HANDBOOK OF ENVIRONMENTAL ENGINEERING™ VOLUME 8

Biological Treatment Processes

Edited by

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Biological Treatment Processes

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Preface

The past thirty years have seen the emergence of a growing desire worldwide that positive actions be taken to restore and protect the environment from the degrading effects of all forms of pollution – air, water, soil, and noise. Since pollution is a direct or indirect consequence of waste, the seemingly idealistic demand for "zero discharge" can be construed as an unrealistic demand for zero waste. However, as long as waste continues to exist, we can only attempt to abate the subsequent pollution by converting it to a less noxious form. Three major questions usually arise when a particular type of pollution has been identified: (1) How serious is the pollution? (2) Is the technology to abate it available? and (3) Do the costs of abatement justify the degree of abatement achieved? This book is one of the volumes of the *Handbook of Environmental Engineering* series. The principal intention of this series is to help readers formulate answers to the last two questions above.

The traditional approach of applying tried-and-true solutions to specific pollution problems has been a major contributing factor to the success of environmental engineering, and has accounted in large measure for the establishment of a "methodology of pollution control." However, the realization of the ever-increasing complexity and interrelated nature of current environmental problems renders it imperative that intelligent planning of pollution abatement systems be undertaken. Prerequisite to such planning is an understanding of the performance, potential, and limitations of the various methods of pollution abatement available for environmental scientists and engineers. In this series of handbooks, we will review at a tutorial level a broad spectrum of engineering systems (processes, operations, and methods) currently being utilized, or of potential utility, for pollution abatement. We believe that the unified interdisciplinary approach presented in these handbooks is a logical step in the evolution of environmental engineering.

Treatment of the various engineering systems presented will show how an engineering formulation of the subject flows naturally from the fundamental principles and theories of chemistry, microbiology, physics, and mathematics. This emphasis on fundamental science recognizes that engineering practice has in recent years become more firmly based on scientific principles rather than on its earlier dependency on empirical accumulation of facts. It is not intended, though, to neglect empiricism where such data lead quickly to the most economic design; certain engineering systems are not readily amenable to fundamental scientific analysis, and in these instances we have resorted to less science in favor of more art and empiricism.

Since an environmental engineer must understand science within the context of application, we first present the development of the scientific basis of a particular subject, followed by exposition of the pertinent design concepts and operations,

and detailed explanations of their applications to environmental quality control or remediation. Throughout the series, methods of practical design and calculation are illustrated by numerical examples. These examples clearly demonstrate how organized, analytical reasoning leads to the most direct and clear solutions. Wherever possible, pertinent cost data have been provided.

Our treatment of pollution-abatement engineering is offered in the belief that the trained engineer should more firmly understand fundamental principles, be more aware of the similarities and/or differences among many of the engineering systems, and exhibit greater flexibility and originality in the definition and innovative solution of environmental pollution problems. In short, the environmental engineer should by conviction and practice be more readily adaptable to change and progress.

Coverage of the unusually broad field of environmental engineering has demanded an expertise that could only be provided through multiple authorships. Each author (or group of authors) was permitted to employ, within reasonable limits, the customary personal style in organizing and presenting a particular subject area; consequently, it has been difficult to treat all subject material in a homogeneous manner. Moreover, owing to limitations of space, some of the authors' favored topics could not be treated in great detail, and many less important topics had to be merely mentioned or commented on briefly. All authors have provided an excellent list of references at the end of each chapter for the benefit of interested readers. As each chapter is meant to be self-contained, some mild repetition among the various texts was unavoidable. In each case, all omissions or repetitions are the responsibility of the editors and not the individual authors. With the current trend toward metrication, the question of using a consistent system of units has been a problem. Wherever possible, the authors have used the British system (fps) along with the metric equivalent (mks, cgs, or SIU) or vice versa. The editors sincerely hope that this duplicity of units' usage will prove to be useful rather than being disruptive to the readers.

The goals of the *Handbook of Environmental Engineering* series are: (1) to cover entire environmental fields, including air and noise pollution control, solid waste processing and resource recovery, physicochemical treatment processes, biological treatment processes, biosolids management, water resources, natural control processes, radioactive waste disposal and thermal pollution control; and (2) to employ a multimedia approach to environmental pollution control since air, water, soil and energy are all interrelated.

As can be seen from the above handbook coverage, no consideration is given to pollution by type of industry, or to the abatement of specific pollutants. Rather, the organization of the handbook series has been based on the three basic forms in which pollutants and waste are manifested: gas, solid, and liquid. In addition, noise pollution control is included in the handbook series.

This particular book Volume 8, *Biological Treatment Processes*, is a sister book to Volume 9, *Advanced Biological Treatment Processes.* Both books have been designed to serve as comprehensive biological treatment textbooks as well as wide-ranging reference books. We hope and expect they will prove of equal high value to advanced

undergraduate and graduate students, to designers of water and wastewater treatment systems, and to scientists and researchers. The editors welcome comments from readers in all of these categories.

This book Volume 8, Biological Treatment Processes, covers the subjects, of fundamental biological concepts, wastewater land application subsurface application, submerged aeration, surface aeration, spray aeration, activated sludge processes, pure oxygen activated sludge process, waste stabilization ponds, lagoons, trickling filters, rotating biological contactors, sequencing bath reactors, oxidation ditch, biological nitrification, denitrification, anaerobic digestion, aerobic digestion, composting, vermicomposting, odor control and VOC control. The sister book Volume 9, Advanced Biological Treatment Processes, covers the subjects of biological process kinetics, vertical shaft bioreactors, aerobic granulation technology, membrane bioreactors, SBR nutrient removal, simultaneous nitrification and denitrification, single-sludge nutrient removal system, nitrogen removal process selection, column bioreactor, upflow sludge blanket filtration, anaerobic lagoons, storage ponds, vertical shaft digestion, flotation, biofiltration, biosolids land application, deep-well injection, natural biological processes, emerging suspended growth biological processes, emerging attached growth biological processes and environmental engineering conversion factors.

The editors are pleased to acknowledge the encouragement and support received from their colleagues and the publisher during the conceptual stages of this endeavor. We wish to thank the contributing authors for their time and effort, and for having patiently borne our reviews and numerous queries and comments. We are very grateful to our respective families for their patience and understanding during some rather trying times. The editors are especially indebted to Dr. Nazih K. Shammas of the Lenox Institute of Water Technology, Massachusetts, for his services as Consulting Editor of this Volume.

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1
Fundamental Concepts for Environmental Processes

Mary Lou Bungay and Henry R. Bungay

CONTENTS

Abstract Living microorganisms consume organic material in wastes, and use its energy to sustain normal activities, to grow, and to reproduce. A biological process, either natural or artificial, involves biochemical reactions, nutrient balance, microbial population balance, and waste disposal. This chapter introduces biological concepts for environmental control processes. The specific topics covered include: cellular interactions, biochemistry, photosynthesis, chemosynthesis, respiration, microbiology, ecology, ecosystem, waste treatment, pollution indices, and biological interactions.

Key Words Biological process • cell • biochemistry • photosynthesis • chemosynthesis • respiration • environmental microbiology • ecology • waste treatment • ecosystems.

1. INTRODUCTION

Sound foundations with understanding of reactions and processes are essential to environmental scientists and engineers for determining the fate of pollutants that reach natural systems and for the improvement of waste treatment. Most of the economically effective methods for destroying wastes use normal cellular processes for breakdown of many types

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of organic wastes. Concepts from biochemistry and biology that are introduced in this chapter apply to specific treatment processes covered in subsequent chapters.

Living cells consume organic material and use its energy to sustain normal activity, to grow, and to reproduce. Some of the cells' wastes—water, carbon dioxide, and minerals—are environmentally acceptable. The cellular mass, however, is itself pollutional because its discharge into streams and lakes would provide nutrients for microorganisms that consume oxygen; thus fish could suffocate. Biological waste treatment usually strives to produce a minimal amount of cellular material that is easily collectible for disposal.

Proper regard for basic scientific principles provides a basis for achieving high efficiency for treatment processes. Although understanding is incomplete because of the great complexity of bioprocesses containing ill-defined nutrients and many different organisms, there have been practical results in terms of design and processing through considering biochemistry and biology.

2. THE CELL

In a scientific context, life is most adequately described in terms of activity. An entity that is organized so as to maintain a definite structure, respond to stimuli, grow, reproduce its own kind, and acquire the energy needed for all of these activities is generally regarded as a living organism. The cell is the structural and functional unit of life. In multicellular organisms, cells are often highly specialized and function in cooperation with other specialized cells. But many organisms are, in fact, free-living single cells.

Although cells differ in size, shape, and specialization, all have the same basic structure. Every cell is composed of *cytoplasm*, a colloidal system of large organic molecules integrated with a complex solution of smaller organic molecules and inorganic salts. The cytoplasm is bounded by a semielastic, selectively permeable *cell membrane* that controls the movement of molecules into and out of the cell. Threadlike *chromosomes* suspended in the cytoplasm bear a linear arrangement of *genes*. Information carried on the genes controls every cellular activity, and, as the units of heredity, genes determine the characteristics of cells from one generation to the next.

In most cells, the chromosomes are surrounded by a cell membrane to form a conspicuous *nucleus*. A number of other organized intracellular structures serve as specialized sites for cellular activities. Certain cells of green plants, for example, contain *chloroplasts* that play an essential role in photosynthesis. Chlorophyll and other associated photosynthetic pigments are contained within the layered membranous structure of the chloroplast. Cells that possess organized nuclei are *eukaryotic*.

In bacteria, archea, and cyanobacteria (formerly called blue-green algae) the chromosomes are not surrounded by a membrane, and there is little apparent subcellular organization. The chlorophyll of cyanobacteria is associated with loosely arranged membranes within the cytoplasm; bacterial chlorophyll, when present, is located in vesicular *chromatophores*. Because they lack a discrete nucleus, these organisms are said to be *prokaryotic*.

Many cells are surrounded by an outer covering external to the cell membrane. Plant cells, bacteria, and blue-green algae are protected by rigid *cell walls*. Certain algae and protozoa are surrounded by siliceous shells.

The distinctive and sometimes elaborate shape exhibited by many unicellular organisms is an inherited characteristic. However, evidence gathered in the culture of isolated cells suggests that in multicellular organisms, cell shape is environmentally determined.

The smallest known cell, pleuropneumonia-like organism (PPLO) is approximately 0.1 micron (μm) in diameter, and the largest, the ostrich egg is about 150 mm in diameter. Most cells, however, have diameters of 0.5 to 40 μm. Because all of the substances required by the cell must enter through the surface membrane, one of the most important limitations to cell size is the ratio of surface to volume. The ease with which a given substance passes through the membrane, its rate of diffusion through the cytoplasm, and the rate at which it is used by the cell have a bearing on cell size. Another important factor in cell size is the proximity of the genes, which continuously monitor cellular activity; as cell size increases, interaction with remote parts of the cell diminishes.

3. BIOCHEMISTRY

3.1. Important Compounds

Despite the obvious diversity of living forms, there is a surprising consistency in the chemical nature of all living things. The main categories of biochemicals in virtually every living system are carbohydrates, lipids, proteins, and nucleic acids.

Carbohydrates are composed of carbon, hydrogen, and oxygen, commonly in a ratio of 1:2:1 ($C_nH_{2n}O_n$). Carbohydrates that will not form simpler compounds upon the addition of water (hydrolysis) are called simple sugars, or *monosaccharides*. Simple sugars contain from three to seven carbons; the most common sugar is glucose, a six-carbon molecule. With the removal of a molecule of water (condensation), two simple sugars may combine to form a *disaccharide*. For example, the disaccharide maltose contains two molecules of glucose (Figure 1.1); the condensation of glucose and fructose, another six-carbon sugar, produces sucrose, or cane sugar.

In the same manner a large number of monosaccharide units may be joined to form polysaccharides such as starch, glycogen, or cellulose (Figure 1.2). Starch and glycogen are energy storage compounds. Cellulose is a major structural material in plants.

Lipids are also made up of carbon, hydrogen, and oxygen. Fats are a very common form of lipid composed of a molecule of glycerol and three fatty acid molecules. Fatty acids are characterized by a long carbon chain and, like all organic acids, by a carboxyl group, $-COOH$. Figure 1.3 shows the general configuration of a *triglyceride* in which R, R′, and R″ represent the carbon chains of three different fatty acids.

Palmitic and oleic acids are examples of two common fatty acids (Figure 1.4). Naturally occurring fats are mixtures of compounds of glycerol with several different fatty acids. Fats serve as storage compounds for reserve energy.

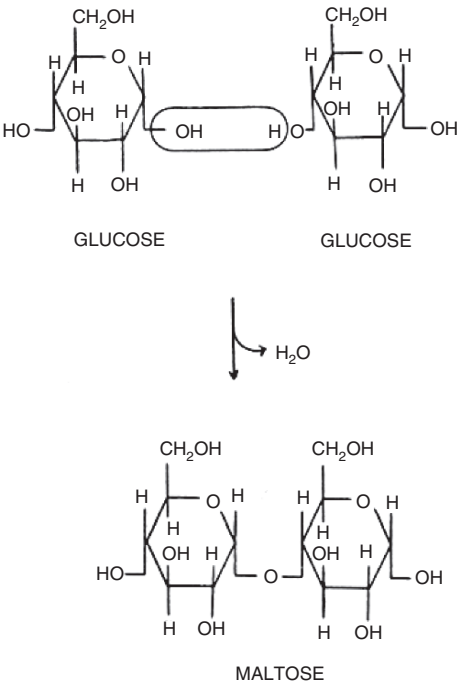

Fig. 1.1. Condensation of two molecules of glucose to form maltose and water.

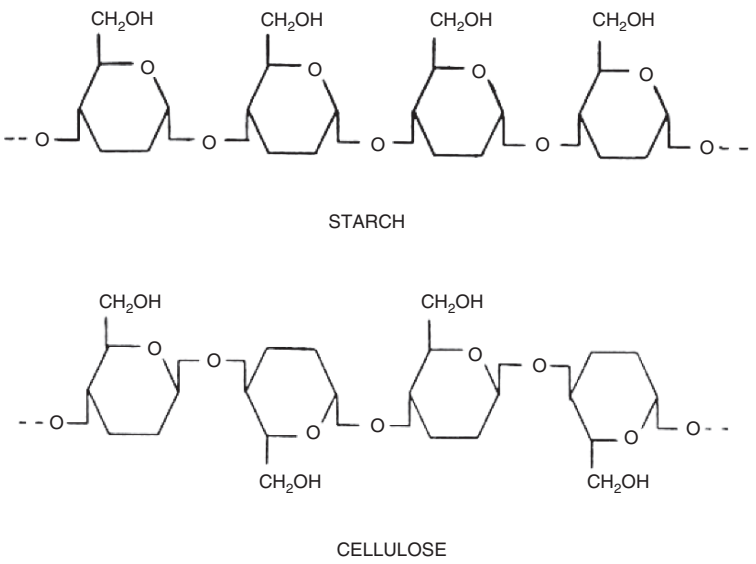

Fig. 1.2. Starch and cellulose.

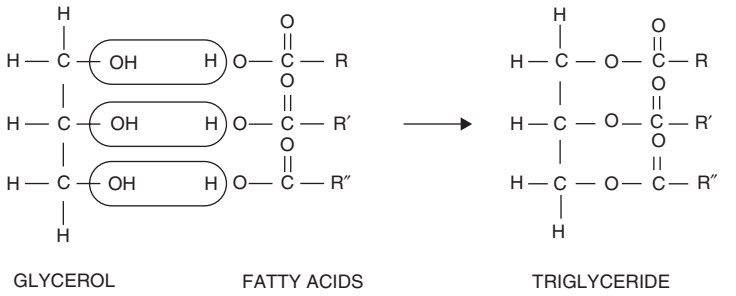

Fig. 1.3. Formation of fats.

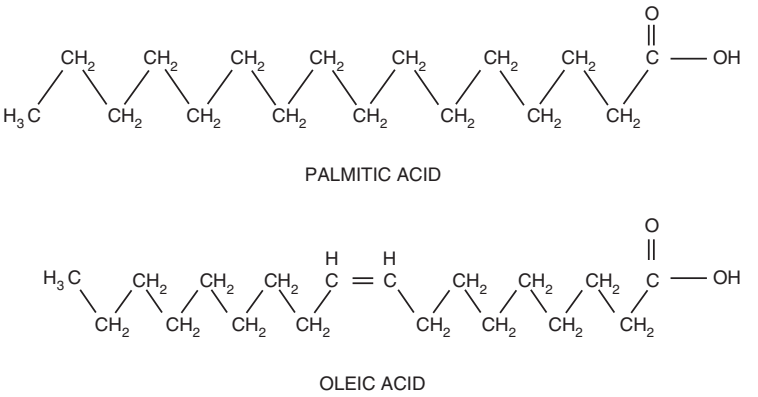

Fig. 1.4. Palmitic acid and oleic acid.

Lipids other than fats found in living systems include phospholipids that play an important role in cell membrane structure and in brain and nerve cells; waxes, that protect the leaves of plants and skins of animals; and steroids, which act as regulatory agents, such as hormones.

Proteins are composed of units called *amino acids*. There are twenty amino acids commonly found in naturally occurring proteins. Amino acids are characterized by a carboxyl group and an amino function such as $-NH_2$. Sulfur is incorporated into the structure of certain amino acids. The general structure and three representative amino acids are shown in Figure 1.5.

With the removal of a molecule of water between the carboxyl group of one and the amino group of the other, two amino acids may be joined by *a peptide bond* to form a *dipeptide*. Several amino acids bonded in this manner form a *polypeptide* (Figure 1.6).

A naturally occurring polypeptide of many amino acids is called a protein. Because of the great length of protein chains, the possible sequences of amino acids, and spatial arrangements, the variety of proteins is essentially infinite. In addition to peptide bonds, other bonds may be formed, giving the molecule a complex and distinctive configuration. In the presence

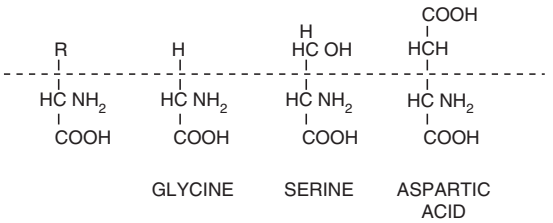

Fig. 1.5. Amino acid structure.

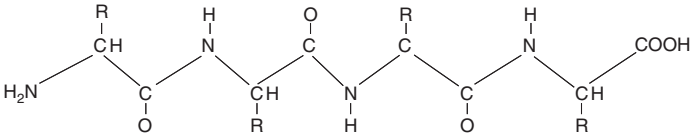

Fig. 1.6. Polypeptide chain.

of certain chemical reagents, excessive heat, radiation, or unfavorable pH, the structure may become disorganized. Proteins are important components of cell membranes and of muscle. The antibodies that protect organisms against invasion by foreign proteins are themselves proteins.

A special class of proteins, the *enzymes*, plays a vital role in all cellular activity. To initiate any chemical reaction, a certain amount of energy is required. Heat could provide the necessary *activation energy* but the amount of heat that would be needed to initiate many biological reactions would destroy the cell itself. Enzymes are the biological catalysts that expedite reactions by lowering the amount of activation energy required.

Virtually every cellular reaction requires the presence of an enzyme. As reactant molecules come into contact with the enzyme surface, an enzyme-substrate complex is formed. When the reaction is complete, the complex dissociates, freeing the enzyme for further reaction. Because of this reuse, only small amounts of enzyme are needed.

The variety and complexity of surfaces of enzymes accounts for their specificity; most enzymes will catalyze only a single reaction or a few closely related reactions. The optimum pH for most enzymes is not far from neutral; most lose activity quickly at temperatures above 60°C.

Enzymes function in conjunction with another special class of compounds known as *coenzymes*. Coenzymes are not proteins; many of the known coenzymes include vitamins, such as niacin and riboflavin, as part of their molecular structure. It is the coenzymes that carry reactant groups or electrons between substrate molecules in the course of a reaction. Because coenzymes serve merely as carriers and are constantly recycled, only small amounts are needed to produce considerable amounts of biochemical product.

Two kinds of *nucleic acids* are found in living organisms: deoxyribonucleic acid (DNA) and ribonucleic acid (RNA). Nucleic acids are chains of *nucleotides*. Each nucleotide consists of a nitrogen-containing organic base, a sugar, and a phosphate group. The sequence in which the

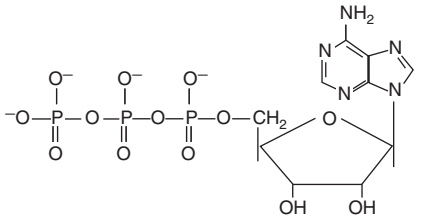

Fig. 1.7. Adenosine triphosphate (ATP).

nucleotides are arranged is actually the code that determines the amino acids to be assembled, and in what order, to form proteins.

The ultimate control of all cellular activity rests with the nucleic acids. Enzymes are required for each cellular reaction, and thus have immediate control. But it is nucleic acids that dictate the synthesis of enzymes. Moreover, it is nucleic acids that are responsible for the maintenance of genetic continuity. When any organism reproduces, equivalent DNA molecules are transferred to each offspring. Even a slight alteration in the nucleotide sequence of a DNA chain may result in some permanent change, or *mutation*, that will persist through succeeding generations.

Each of the activities implicit in the word "life" requires energy. In living things energy is stored and transferred as chemical bond energy. The multitude of reactions that takes place within a living system is collectively termed *metabolism*.

Certain metabolic reactions, once activated, proceed spontaneously with a net release of energy. Hydrolysis and molecular rearrangements are examples of spontaneous reactions. The hydrolytic splitting of starch to glucose, for instance, results in a net release of energy. The energy can be made available for a different reaction.

A great many biochemical reactions are not spontaneous and therefore require energy input. In living systems this requirement is met by coupling an energy-requiring reaction with an energy-releasing reaction. The synthesis and breakdown of biochemical compounds is achieved through pathways involving the formation of energy-rich intermediate compounds. In this way energy can be transferred in a stepwise manner.

If a sufficient amount of energy is produced by a metabolic reaction, it may be used to synthesize a high-energy compound. Adenosine triphosphate (ATP) is such a compound. ATP is a nucleotide consisting of the nitrogen-containing compound, adenine, the sugar ribose, and three phosphate groups (Figure 1.7).

Although ATP has adequate stability for the short-term, it hydrolyzes spontaneously in water. When the terminal phosphate linkage is broken, adenosine diphosphate and inorganic phosphate are formed, and energy is provided. When sufficient energy becomes available, ATP can be reformed.

Oxidation and reduction are very common steps in metabolism. Reduction reactions store energy in the reduced compound, whereas oxidation liberates energy. In biological systems, the most frequent mechanism of oxidation is the removal of hydrogen, and conversely, the addition of hydrogen is the most frequent method of reduction. When this takes place

within a cell, hydrogen is transported between donor and acceptor molecules by coenzymes. Nicotinamide-adenine dinucleotide (NAD) and nicotinamide-adenine dinucleotide phosphate (NADP) are two coenzymes that function in this manner.

3.2. Photosynthesis

Every living thing can synthesize ATP, but only green plants and a few microorganisms have the capacity to make it from energy-poor materials. Through the process of *photosynthesis*, these organisms are able to convert light energy to chemical bond energy and reduce carbon dioxide to carbohydrate.

When light strikes a photosynthetic organism, energy is absorbed by an array of pigments including chlorophyll. This energy is used to convert ADP to ATP and to reduce NADP by the addition of hydrogen ions donated by water.

$$2\text{ADP} + \text{P}_\text{i} + 2\text{NADP}^+ + 4\text{H}_2\text{O} + \text{light energy} \rightarrow 2\text{ATP} + 2\text{NADPH} + \text{O}_2 + 2\text{H}_2\text{O}$$

Because light is essential for the production of ATP and the reduction of NADP, these events are known as the *light reactions* of photosynthesis. The light reactions in photosynthetic bacteria differ somewhat from the green plants. Bacteria do not use water as a source of hydrogen ions and oxygen is not formed. Some use organic molecules, others use hydrogen sulfide (H_2S) and give off sulfur.

The remaining reactions can take place whenever ATP, NADPH, and carbon dioxide are present; they are, therefore, called the *dark reactions*. In these reactions, CO_2 combines with a five-carbon sugar that immediately splits to form two molecules of a three-carbon compound, phosphoglycerate (PGA), and PGA is reduced to phosphoglyceraldehyde (PGAL). Five-sixths of the PGAL is used to regenerate the five-carbon sugar, ribulose diphosphate, through a complicated series of reactions, and it once again combines with CO_2. The remainder of the PGAL is used in the synthesis of sugars and starch. The dark reactions of photosynthesis are summarized in Figure 1.8.

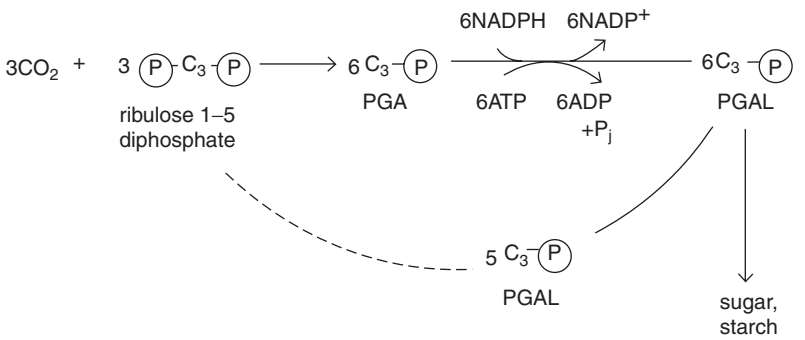

Fig. 1.8. Dark reactions of photosynthesis.

3.3. Chemosynthesis

The ultimate source of energy for most living things is the sun. But certain groups of bacteria require neither light nor organic energy sources. These organisms derive energy from the oxidation of inorganic substances, and are called the *chemosynthetic* bacteria.

For example, one species of nitrifying bacteria oxidizes ammonia to nitrite, and another species oxidizes nitrate to nitrate:

$$2NH_3 + 3O_2 \rightarrow 2HNO_2 + 2H_2O + \text{energy}$$
$$HNO_2 + {}^1/_2\, O_2 \rightarrow HNO_3 + \text{energy}$$

Certain microorganisms oxidize elemental sulfur to sulfate:

$$2S + 3O_2 + 2H_2O \rightarrow 2H_2SO_4 + \text{energy}$$

Species of archea oxidize hydrogen gas, reducing carbon dioxide to methane:

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O + \text{energy}$$

3.4. Respiration

The oxidative breakdown of organic molecules is called *respiration*. It is through this process that the cell recovers energy stored in organic substances. Respiration is really a controlled series of dehydrogenations in which small amounts of energy are released at several stages. The released energy is incorporated into ATP, where it is readily available for other reactions. As in all metabolic pathways, a specific enzyme is required at each step.

The first stage in the breakdown of carbohydrate is the same in all organisms. Glucose is oxidized to form two molecules of the three-carbon compound pyruvic acid. The series of eight reactions, termed *glycolysis*, is outlined in Figure 1.9.

In many organisms, including man, respiration can proceed only in the presence of molecular oxygen (*aerobic* respiration). There are organisms, however, that can carry on respiration in the absence of oxygen (*anaerobic* respiration). Anaerobic respiration occurs in many microorganisms and, under certain conditions, in the muscle cells of animals. Most bacteria are *facultative anaerobes*, growing in the presence or absence of oxygen. Some microorganisms, the *obligate anaerobes*, require the absence of oxygen. *Obligate aerobes*, on the other hand, must have molecular oxygen.

The aerobic oxidation of pyruvic acid is outlined in Figure 1.10.

Carbon dioxide is removed from pyruvic acid, leaving a two-carbon acetate group. Acetate is carried by a coenzyme (Coenzyme A, or CoA) into the *citric acid cycle*. No oxygen is taken up in the citric acid cycle, but a series of oxidations takes place in which hydrogen is transferred to coenzymes, and further removal of carbon dioxide occurs. The hydrogen is passed from the coenzymes through a series of carrier molecules called the respiratory chain, or cytochrome system, and finally to oxygen. Energy produced during the reduction of the cytochrome molecules is used to convert ADP to ATP. Oxygen is required only as the final hydrogen acceptor.

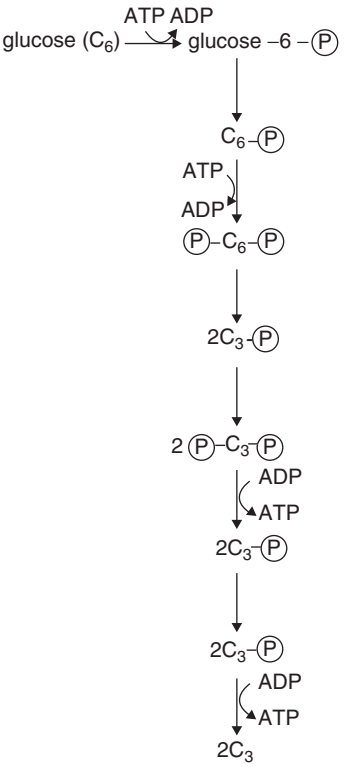

Fig. 1.9. Glycolysis.

Aerobic respiration can be represented by the equation:

$$C_aH_bO_c \cdot N_dP_e \cdot S_f + n\,O_2 \rightarrow m\,CO_2 + p\,H_2O + q\,NH_3 + r\,PO_4^{3-} + s\,SO_4^{2-}$$

The coefficients of the products are related stochiometrically to the subscripts of the reactants, but the equation is simplistic. In reality the subscripts vary from one type of organic compound to another and additional atoms are sometimes present. The products can include mixtures of organic compounds of varying oxidation states, as well as inorganic ions other than those shown.

In anaerobic respiration, molecules other than oxygen act as hydrogen acceptors. In many cases pyruvic acid itself accepts the hydrogen liberated during glycolysis. This process, called *fermentation*, produces either lactic acid, or a two-carbon alcohol plus carbon dioxide.

Glycolysis produces a net gain of two ATP molecules. Further anaerobic oxidation of pyruvic acid adds none, thus anaerobic respiration of one molecule of glucose yields only two ATP. On the other hand, aerobic oxidation of one glucose to carbon dioxide and water, via glycolysis and the citric acid cycle, produces a total of 38 new molecules of ATP.

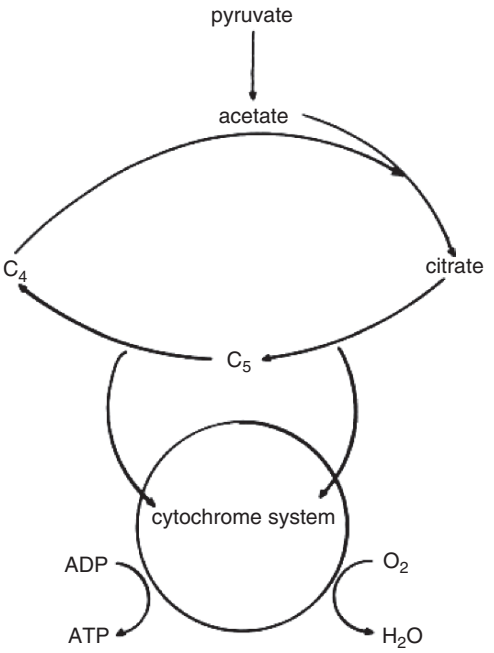

Fig. 1.10. Citric acid cycle.

In addition to carbohydrates, cells regularly oxidize fats as a source of energy. Proteins and amino acids are less frequently broken down, but under starvation conditions, they, too, may be oxidized to provide needed energy.

3.5. Nutrition

All organisms require sources of carbon, nitrogen, sulfur, phosphorus, water, and certain trace elements. Some have specific vitamin requirements as well. Water, in addition to its role as a reactant and as the principal component of cytoplasm, serves as the medium through which molecules are transferred into and out of the cell. All cells, even those of terrestrial organisms, require constant moisture to remain active.

Green plants need only carbon dioxide, nitrate or ammonium ions, dissolved minerals, and water to manufacture all of their cellular components. Photosynthetic bacteria require an additional specific source of hydrogen ions, and the chemosynthetic bacteria must have a specific oxidizable substrate. Some microorganisms have the ability to "fix" atmospheric nitrogen by reducing it to ammonia. Organisms that use only simple inorganic compounds as nutrients are said to be *autotrophic* (self-nourishing).

Organisms that require compounds that have been manufactured by other organisms are called *heterotrophs* (other-nourishing). Heterotrophs such as the fungi, which use only dead material, are known as *saprophytes*, those that live in or on other living organisms using compounds produced by the host are *parasites*.

Because organic molecules frequently are too large or insoluble to pass through the cell membrane, many heterotrophs produce enzymes that act outside the cell. These *exoenzymes* hydrolyze large molecules to smaller units that can readily enter the cell.

4. MICROBIOLOGY

Every living thing can be assigned to one of three *domains*: Bacteria, Archaea, or Eukarya. Bacteria, Archaea, and certain eukaryotes—protozoa, algae and fungi—are regularly represented in waste treatment systems. Viruses are also present, and some infect microorganisms. Biologists classify living things according to their derivation from common ancestors. Among microorganisms, however, the relationships are seldom clear-cut so that classification is often arbitrary and confusing.

4.1. Bacteria

Bacteria are tiny single cell organisms ranging from 0.5 to 20 μm in size, although some may be smaller and a few exceed 100 μm in length. The typical bacterial cell is surrounded by a rigid protective membrane made up of compounds called mucoproteins that are peculiar to bacteria. It is this wall that imparts characteristic shape to the cell: round or ovoid, rod-shaped, or spiral. Most bacteria are hyperosmotic relative to their surrounding medium; it is the rigid cell wall that prevents their swelling and bursting.

Bacteria are prokaryotic, which means they possess no organized nuclei or organelles; they do, of course, contain genetic material, both DNA and RNA, and the cytoplasm may contain numerous granules composed of carbohydrates, fats, and other nutrients. When chlorophyll is present, it is of a type unique to bacteria. Many bacteria exhibit motility by means of one or more hairlike appendages called flagella. Flagella are not uncommon in other types of organisms as well, but the microstructure of the bacterial flagellum is, again, unique. Bacteria reproduce by dividing into parts (usually equal), a process termed binary fission.

Under unfavorable conditions, certain bacteria can transform to spores that germinate upon return to a favorable environment. Many species of bacteria may, under appropriate conditions, become surrounded by a gelatinous material. If a number of cells share the same gelatinous mass, it is called a slime; if the cells are separately surrounded, each is said to have a capsule. The slime or capsule affords the cell a means of attachment and provides a measure of protection against drying and predators.

Bacteria are classified on the basis of pathogenicity, morphology, and physiological characteristics; some are also characterized by the arrangement of cells in clusters, chains, or discrete packets. They cover the entire spectrum of nutritional requirements from photosynthetic autotrophs to the most fastidious of heterotrophs. Many possess exocellular enzymes that allow them to break down a variety of complex substrates to molecules that can enter the cell to be further metabolized. Each species of bacteria grows best within certain ranges of pH and temperature, commonly not far from neutral pH and between 25 and 40°C. Those that grow best in this temperature range are called *mesophiles*. Bacteria that achieve maximum growth below 20°C are called *psychrophiles*, and those favored by temperatures above 45°C are designated *thermophiles*. Many bacteria synthesize pigments that impart distinctive colors

to their colonies. Using special staining techniques, individual cells of most species are visible with the light microscope; some, however, are best observed by electron microscopy.

4.2. Archaea

Archaea were long considered bacteria, but recent discoveries have led to placement in a separate domain. Like bacteria, they are single-cell organisms with no organized nucleus. However, at the molecular level they are quite different in, for example, the composition of their membranes, cell walls, and RNA. Archaea are often found in environments of extreme temperature, pH, or salinity. Some archaea can grow at 113°C. Archaea of importance in waste treatment are the anaerobic *methanogens*, which produce methane gas.

4.3. Algae

Algae are aquatic organisms containing photosynthetic pigments that enable them to synthesize structural materials and storage compounds from carbon dioxide and water. The cellular activities of algae significantly affect the oxygen resources of surface waters. The distinctive colors imparted by their pigments are one of the criteria by which algae are classified. Some species are unicellular and microscopic; others are filamentous, branched, or colonial. Some have life cycles in which both unicellular and multicellular forms arise, but the most common mode of reproduction is simple cell division. On the basis of pigmentation, storage compounds, cell organization, and morphology, biologists divide the algae into as many as nine groups. There are three groups of fresh-water algae.

Cyanobacteria (blue-green algae) may be unicellular, colonial, branched or filamentous. The photosynthetic pigments are not organized into discrete structures (chloroplasts or chromophores), but are dispersed throughout the cytoplasm. The cell wall is usually very thin and may be enclosed within a gelatinous sheath. Cyanobacteria do not contain starch; their storage compounds are glycogen-like substances. Like the bacteria, they possess no distinct nucleus and are, therefore, prokaryotic. Also, like some bacteria, Cyanobacteria have the ability to convert atmospheric nitrogen to ammonia (nitrogen fixation) that can be used in the synthesis of organic compounds or excreted into the medium.

Chlorophyta (green algae) may be free-swimming or attached. The cells are eukaryotic; that is, each has a distinct nucleus and chlorophyll is contained in chloroplasts. Starch is the predominant storage compound. Individuals may be branched, filamentous, colonial, or single cells; often they are microscopic, but may become so numerous as to be visible as an algal "bloom" or scum on the surface of standing water.

Chrysophyta (yellow-green or yellow-brown algae) are unicellular or colonial. All species are motile and surrounded by a thick cell wall; in some forms (the Diatoms) the wall is impregnated with silicon. Starch is not present and food is stored as lipids, which often gives members of Chrysophyta a metallic luster.

4.4. Protozoa

The protozoa are a widely diverse group of organisms of 15,000 to 20,000 known species. Most are microscopic, although some attain a length as great as 5 mm. A cell membrane encloses the cytoplasm, and within the cytoplasm are found a number of cellular inclusions,

or organelles, that are the sites of specialized cell functions. For this reason, protozoa are often referred to as *acellular* rather than unicellular. In addition, many forms are truly multicellular during certain stages of their life cycles and some contain multiple nuclei throughout their lives. Some protozoa are surrounded by cell walls or a shelllike covering, others are not; the majority of species are individuals, but some are colonial; most are freeliving and actively motile, but a few species remain attached to surfaces throughout their adult lives. Those species that contain chlorophyll are regarded by many botanists as algae.

The protozoa are divided into four groups on the basis of morphology: Mastigophora, Ciliophora, Sarcodina, and Sporozoa. The Mastigophora possess one or more whiplike appendages, called *flagellae*. Ciliophora have numerous, shorter, hairlike appendages known as *cilia*. The Sarcodina have neither flagellae nor cilia, but move and engulf food particles by constantly changing extensions of the protoplasm called *pseudopodia*. All of the Sporozoa are parasitic and have complicated life cycles. Species of Sporozoa are the agents of such diseases as malaria and coccidiosis.

The most common method of reproduction among the protozoa is binary fission. Nutritionally, protozoa range from the photosynthetic autotrophs to the parasites. Heterotrophic forms ingest small food particles such as bacteria, other protozoa, or even small invertebrates. The food is digested within the cytoplasm to compounds that can be metabolized by the organism.

4.5. Fungi

As a group, fungi have simple vegetative bodies from which reproductive structures are elaborated. All fungal cells possess distinct nuclei and, at some stage in their life cycles, reproduce by spores formed in specialized fruiting bodies. The fungi contain no chlorophyll and therefore require sources of complex organic molecules; many species grow on dead organic material, others live as parasites. Many can live on carbohydrate, inorganic nitrogen, and salts. Food is stored as glycogen or oil.

Fungi are classified as slime molds, true fungi, or yeasts, based upon vegetative and reproductive structure. The somatic (vegetative) stage of the "true" slime molds is a multinucleate amoeboid mass, generally 2 to 3 in. in length, called a *plasmodium*. The entire plasmodium moves about engulfing food particles, but under certain conditions, it becomes stationary and develops fruiting bodies that produce spores. Products of spore germination fuse, divide, and grow forming a new plasmodium. A second kind of slime mold, the cellular slime mold is an aggregate of many individual amoeboid forms. The plasmodium is formed only when individual cells fail to find sufficient food. The "pseudoplasmodium;" or mass of individual cells, becomes stationary and fruiting bodies develop, forming spores that germinate as individual amoeboid cells.

Molds, mildews, and mushrooms are true fungi. The vegetative body, or *thallus*, of a true fungus consists of elongated filamentous structures called *hyphae*, and a mass of hyphae is called a mycelium. The mycelia of some fungi are distinctively colored, for example, the black bread mold *Rhizopus* and green mold *Penicillium*. Specialized hyphae anchor the mycelium to its substrate, and others become reproductive bodies that produce spores. Each spore may become a new mycelium. Fungi are often indistinguishable in their vegetative stages, and

are classified on the basis of fruiting bodies. Mushrooms and toadstools are among the most conspicuous fungal fruiting bodies.

Yeasts are nonfilamentous fungi and, therefore, do not form mycelia. They are unicellular organisms surrounded by a cell wall and possessing a distinct nucleus. Most yeasts reproduce by a process known as budding; a small new cell is pinched off the parent cell, but under certain conditions an individual yeast cell may become a fruiting body, producing four spores. The spores are more resistant than vegetative cells to extremes of temperature and prolonged periods of drying, enabling yeasts to survive unfavorable environmental conditions.

4.6. Viruses

Viruses are particles too small to be seen with a light microscope. They are not cellular in structure and are composed mainly of nucleic acid polymers surrounded by a protein sheath. Lacking metabolic machinery, viruses exist only as parasites that replicate within a living cell and are released when cells die and disintegrate. They are highly host specific, infecting only a single species or closely related species. Plant and animal viruses are generally named for the diseases they cause, such as tobacco mosaic virus or influenza virus.

Not all types of microorganisms appear to be susceptible, but bacteria and certain molds are subject to invasion by virus particles. Those that attack bacteria are called *bacteriophages*, and may be either *virulent* or *temperate*. Virulent bacteriophages divert the cellular resources to the manufacture of phage particles and kill the cell. Temperate bacteriophages have no immediate effect upon the host cell; they become attached to the bacterial chromosome and may be carried through many generations before being triggered to virulence by some physical or chemical event.

4.7. Other

Flatworms, roundworms, rotifers, insects, insect larvae, and tiny crustaceans have been identified in wastewater. They are present in small numbers and play a minor role in sewage treatment processes.

Organisms produced by genetic engineering (use of recombinant DNA techniques whereby sections of genetic material are incorporated into that of another organism) have been designed to improve the performance of processes for biological waste treatment. These and other organisms that have been selected because of desired metabolic properties have not yet shown the ability to persist in nature or in the waste treatment processes. This should be expected because natural selection favors those organisms that compete well and adapt to the processes. Organisms that do what humans desire are unlikely to have the characteristics best suited to survival in competition with those selected naturally.

5. ECOLOGY

Clearly, all biological activity is subject to environmental limitations. Physical factors, such as concentrations of dissolved organic and inorganic substances, solar radiation, pH, oxidation-reduction potential, and temperature, impose pressures that determine the selection

of organisms. Aerobic respiration, for example, can only take place where oxygen is present, and aerobes will prosper whereas anaerobes decline.

Biologists use the term *ecosystem* for a physical environment of specified dimensions, along with all the organisms that occupy it. In ecological parlance, an organism's *habitat* is the place where it lives, and its *niche* the role that it plays in the ecosystem. The word *population* is used to describe a group of individuals of the same species and *population density* is the number of such individuals per unit volume or area. All of the populations inhabiting a specific area constitute a *community*.

5.1. Structure of the Ecosystem

The nutritional and energy relationships within an ecosystem are expressed in *trophic levels*. Autotrophs, requiring only light and simple inorganic substances, are *producers*. Heterotrophs that require substances manufactured by autotrophs are *primary consumers*, and those that depend upon other heterotrophs are *secondary consumers*. The saprophytes, organisms of decay, are the *decomposers* that return dead material to simple molecular form. Figure 1.11 depicts trophic levels.

Nutritional relationships are often described as *a food chain*. This is, however, a deceptively simple description; the situation can become very complex when several species are involved and is more accurately called a "food web."

The flow of energy through an ecosystem is depicted as an *energy pyramid* (Figure 1.12).

The triangular configuration illustrates the diminishment of available energy through successive trophic levels. This is illustrated by waste treatment processes in which tiny organisms are present in tremendous numbers whereas there are relatively few protozoa. Nevertheless,

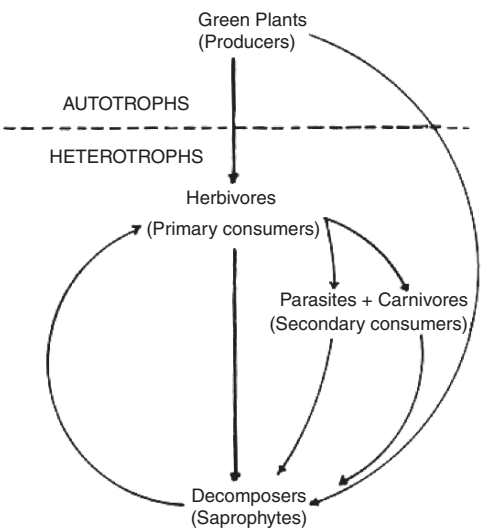

Fig. 1.11. Trophic levels.

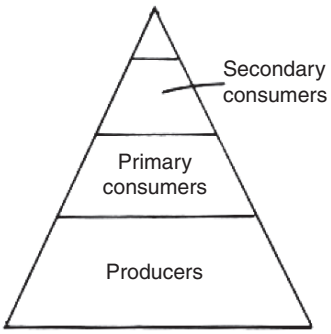

Fig. 1.12. Pyramid of energy.

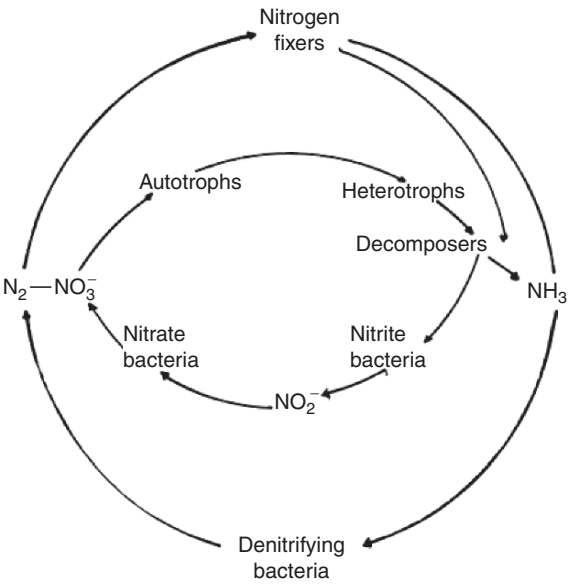

Fig. 1.13. Nitrogen cycle.

these larger creatures through their feeding habits influence the numbers and types of the small organisms.

5.2. Biogeochemical Cycles

Living things require an abundance of carbon, hydrogen, oxygen, and nitrogen, and 30 to 40 other elements in smaller amounts. Each of these elements circulates through the physical and biological components of the environment in a biogeochemical cycle. The familiar nitrogen cycle (Figure 1.13) and the phosphorus cycle (Figure 1.14) are examples.

Figure 1.15 and Figure 1.16 illustrate the carbon and sulfur cycles.

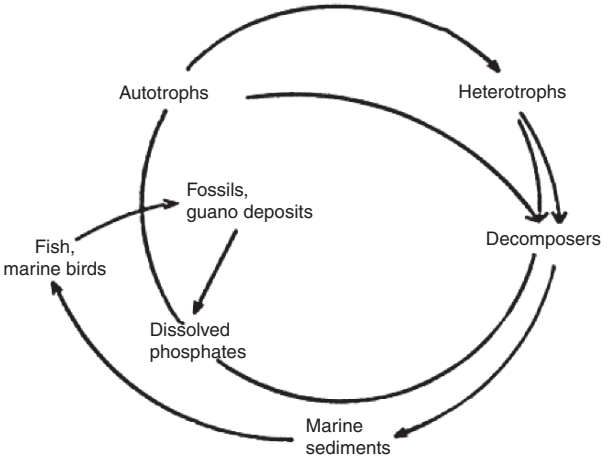

Fig. 1.14. Phosphorus cycle.

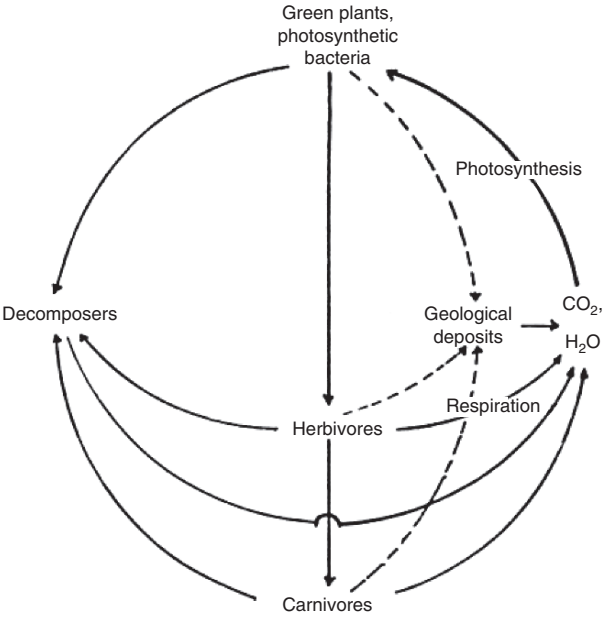

Fig. 1.15. Carbon cycle.

5.3. Interspecies Relationships

In an environment that supports several species, a variety of biological interactions may occur. Mixed-species phenomena are not merely composites of the organisms' pure culture behavior. Biological activity within a complex ecosystem depends upon interactions between

Table 1.1
Common terms for cellular interactions

Term	Type of interaction
Neutralism	Lack of interaction
Commensalism	One member benefits while the other is unaffected
Mutualism	Each member benefits from the other
Competition	A race for nutrients and space
Amensalism	One adversely changes the environment for the other
Parasitism	One depends upon and harms another
Predation	One organism ingests another
Synergism	Cooperative metabolism to produce a substance not produced by either alone

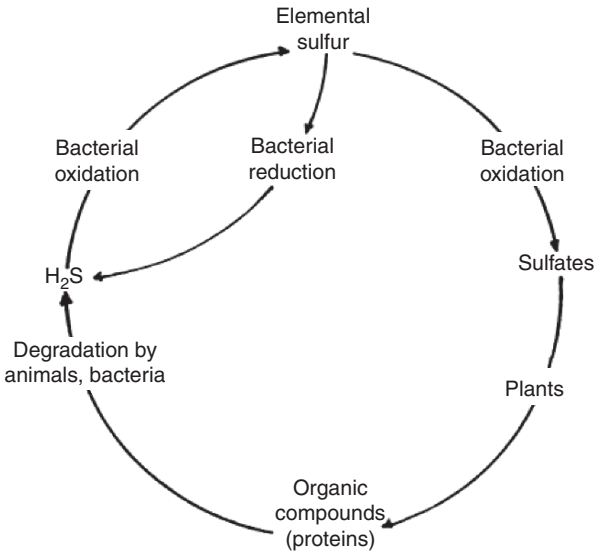

Fig. 1.16. Sulfur cycle.

organisms and upon reciprocal effects between organisms and their environment. Several kinds of interactions are described in Table 1.1.

In a multispecies system, several interactions may take place simultaneously. Moreover, the nature of a relationship may change; for example, one species may stimulate another to the extent that it becomes a competitor.

5.4. Population Dynamics

Theoretically, given ideal conditions, a population of any species would increase indefinitely at an exponential rate. "Ideal conditions" for limitless growth include maximum reproduction, minimum mortality, unlimited resources, and no adverse action by other organisms. In nature this combination does not exist, and natural populations follow a growth pattern

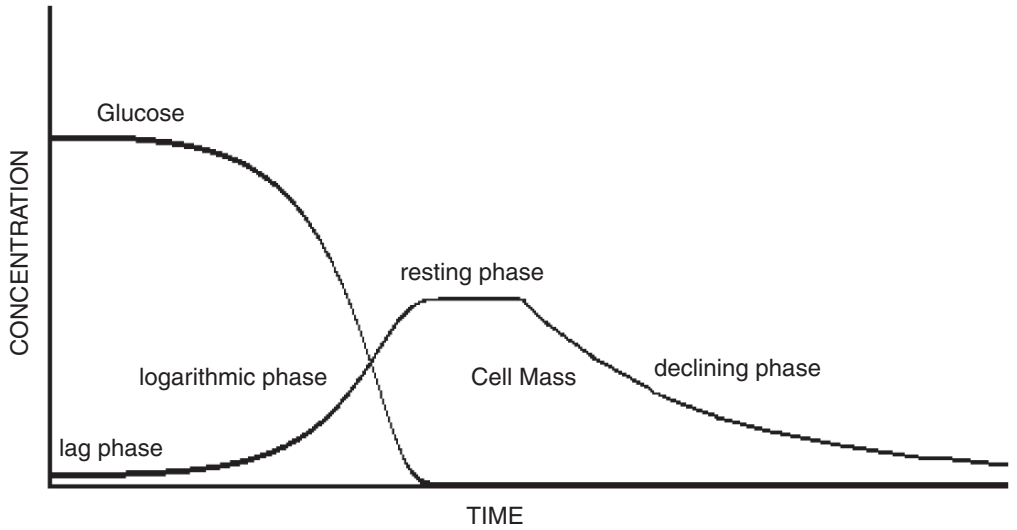

Fig. 1.17. Computer simulation of microbial growth.

similar to that in Figure 1.17, drawn by an internet exercise at http://www.rpi.edu/dept/chem-eng/Biotech-Environ/FUNDAMNT/micro.htm.

During the initial stage, as the population is becoming established, the rate of growth is slow. This is followed by a stage of rapid increase, approaching the maximum rate. Then, as limiting factors come into play, growth slows and the population levels off. Depending upon the limiting factors, it may remain level, decline, or fluctuate.

Population growth of organisms having a very short generation time, such as bacteria, is expressed as the logarithm of the number of individuals versus time. The nature of a

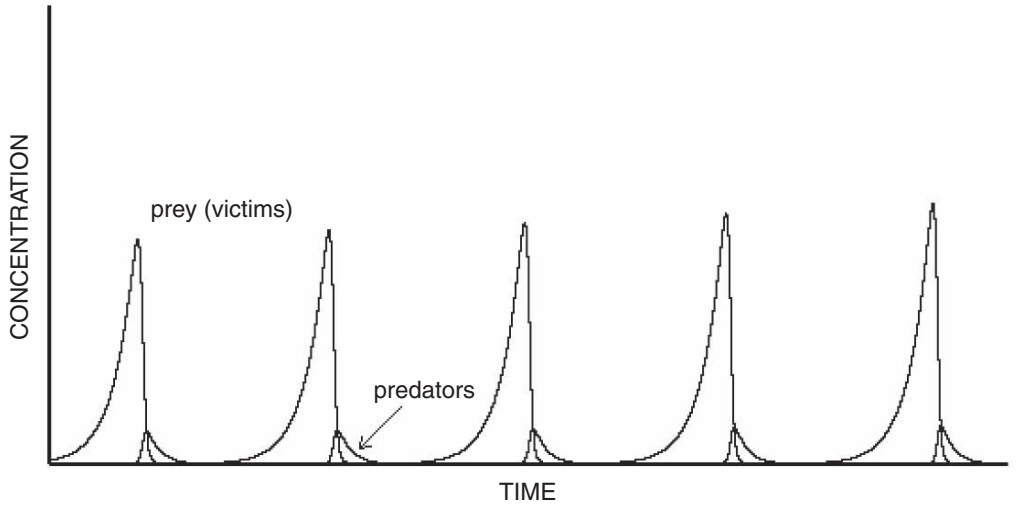

Fig. 1.18. Computer simulation of a prey-predator relationship.

limiting factor can sometimes be inferred from the growth curve. For example, a precipitous decline before attaining maximum population is caused by adverse changes in the physical environment. Some limiting factors are functions of the numbers of organisms, and are said, therefore, to be *density-dependent*. Predation is a density-dependent factor; as the density of prey increases, the density of predator also increases, causing a decline in prey. The decline in prey is followed by a decline in predator population. Figure 1.18 shows fluctuating prey and predator populations.

This figure is drawn by an internet exercise at http://www.rpi.edu/dept/chem-eng/Biotech-Environ/MixCul/predat.html. Competition for environmental resources is also a major density-dependent limiting factor.

6. PHYSICAL AND BIOLOGICAL FACTORS IN WASTE TREATMENT ECOSYSTEMS

6.1. Chemical Composition of the Medium

The chemical composition of the medium is a major factor in determining which species will predominate in any ecosystem. Local collection methods, diet, debris, weather, and other conditions may cause regional differences in the composition of domestic sewage. In general, however, domestic wastes are fairly similar in chemical composition. More than 99 percent is water; the remainder contains an abundance of nitrogenous compounds, with lipids present in lesser amounts. Low molecular weight carbohydrates are in very low concentration, but cellulose often provides a source of organic carbon. Vitamins, minerals, and trace elements are present in adequate amounts for microbial growth. On the whole, domestic waste provides a well-balanced medium for microorganisms, but will favor those that can efficiently use nitrogenous compounds.

Even in a nutritionally complex medium, however, it is extremely unlikely that nutrients will occur in the exact proportions required for the growth of any specific organism. The approximate empirical formula of cellular material is $C_{10}H_{19}O_3N$. If the nutrient medium has proportions other than these, some substance, such as the carbon, nitrogen, or sulfur source, will be in short supply relative to the others. This substance will be the first to be exhausted, and is therefore termed the *limiting nutrient*. If the concentration of this substance should be increased sufficiently, a different nutrient would become growth-limiting. Because nutritional requirements vary, one species may be carbon limited, for example, in the same medium in which another is vitamin or phosphorus limited.

Organisms require a certain concentration of a nutrient to function minimally, without growth. This low level of activity is termed *endogenous metabolism*. The energy required for endogenous metabolism is *maintenance energy*. Figure 1.19 shows growth rate as a function of concentration of limiting nutrient.

The relationship in the figure can be expressed by the equation:

$$\mu = \mu_m \frac{C - C_e}{K - C - C_e}$$

where μ = growth rate coefficient; μ_m = maximum growth coefficient; K = a constant; C = concentration of nutrient; C_e = maintenance concentration.

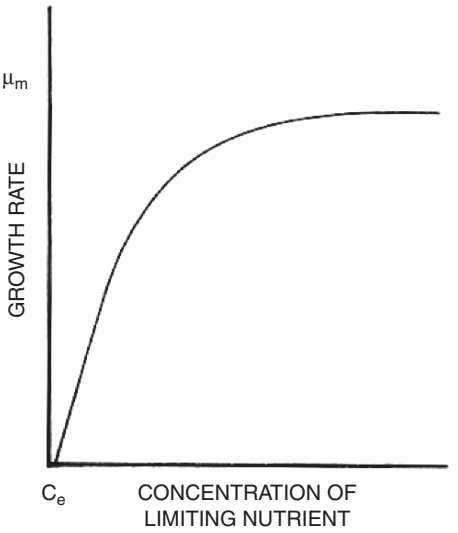

Fig. 1.19. Growth rate as a function of limited nutrient.

6.2. Indices of Pollution

Biological phenomena can reflect pollution of natural waters and the amount of organic matter present in sewage. Domestic water supplies and recreational bathing areas are routinely monitored for the presence of certain microorganisms that indicate contamination. These indicator organisms are members of a group of bacterial species known as coliforms (e.g., *Escherichia coli, Aerobacter aerogenes*) and are always present in large numbers in the intestines of humans and other animals. The presence of such bacteria is taken as an index of fecal pollution and, because the intestines may also harbor pathogenic organisms, the presence of coliforms indicates that the waters are subject to potentially dangerous contamination. Waters may contain other bacterial species that, when particular substrates are present in sufficient amounts, produce slimes, sulfuric acid, or hydrogen sulfide, or cause iron to precipitate. These types of nuisance bacteria can be identified by special tests.

Wastewater contains complicated mixtures of organic materials. Complete identification of all the constituents would be both costly and time-consuming. Consequently, some rather crude methods of measuring organic content are routinely performed. One such method, determination of *biochemical oxygen demand* (BOD), is based upon microbial respiration. To find the BOD, a mixture of microorganisms is allowed to oxidize the organic material under controlled conditions. Oxygen content of the sample is measured before and after a 5-day incubation period and the difference assumed to be the amount used in microbial oxidation of organic molecules. BOD also provides a valuable index of oxygen depletion in waters receiving the organic material.

The organic content of sewage may also be measured by chemical means. *Chemical oxygen demand* (COD) is found by using inorganic chemicals to oxidize organic material. The COD test requires less time, but correlates less well with natural conditions when organic matter

reaches receiving waters. A second chemical determination, *total organic carbon* (TOC) is quick and convenient. In this test organic molecules are chemically oxidized and the amount of carbon dioxide produced is equated to moles of total organic content.

6.3. Flow Rates and Concentration

The typical treatment plant has a continuous but variable input of wastewater. Process loading can be of two types: hydraulic and organic. A disproportionate hydraulic increase in loading occurs when rainwater enters a system of combined sanitary and storm sewers. Increased hydraulic loading dilutes the medium, with the likelihood of reducing metabolic rates, and shortens detention time.

Increases in organic loading are more serious in some types of processes than others. For example, a sudden increase in some toxic component may be devastating to populations in an activated sludge system, whereas organisms embedded in the attached slime layers of a trickling filter are afforded some protection.

Occasionally a waste treatment process is run as a batch operation. A canning plant, for instance, in use for a short time at the end of growing season, might produce a relatively large quantity of waste to be dumped into a lagoon. This kind of operation results in much greater loadings than those of a typical continuous process, but, in compensation, detention time is longer.

6.4. Surfaces and Substrata

The microorganisms in waste treatment processes are either suspended in the medium or attached to surfaces in films or slimes. The availability of attachment sites is, therefore, a factor in species selection. In open tanks and basins, organisms are suspended as single individuals, aggregates, or flocs of many individuals in a common matrix. Attached growth is negligible, although bits of debris may be slime covered. Because individuals and small aggregates settle slowly, they may escape sedimentation. Slimes, flocs, and large aggregates that settle readily are more desirable in waste treatment processes.

Trickling filters, with large expanses of solid surface, favor microbial slimes and films. Some processes, such as rotating disc aerators and oxidation ditches with brush aerators, have significant numbers of both suspended and attached organisms.

6.5. Nutritional Shifts

A mixture of many different organisms growing on many different nutrients can have complicated responses to changes in the feed stream. Some ingredients are used preferentially. For example, feeding a mixture of glucose and lactose (milk sugar), a disaccharide sugar, will give a growth curve as in Figure 1.20.

This figure is drawn by an exercise on the internet at: http://www.rpi.edu/dept/chem-eng/Biotech-Environ/MICROBIOL/diauxie.html. Analysis of the medium will show constant lactose concentration until glucose has declined to a low level. Glucose inhibits synthesis or activation of the enzyme for attacking lactose, and time is required to initiate lactose metabolism after glucose concentration is too low to be inhibitory. The phenomenon of a two-stage growth curve is termed *diauxie*. In mixed cultures that are well-acclimated to both

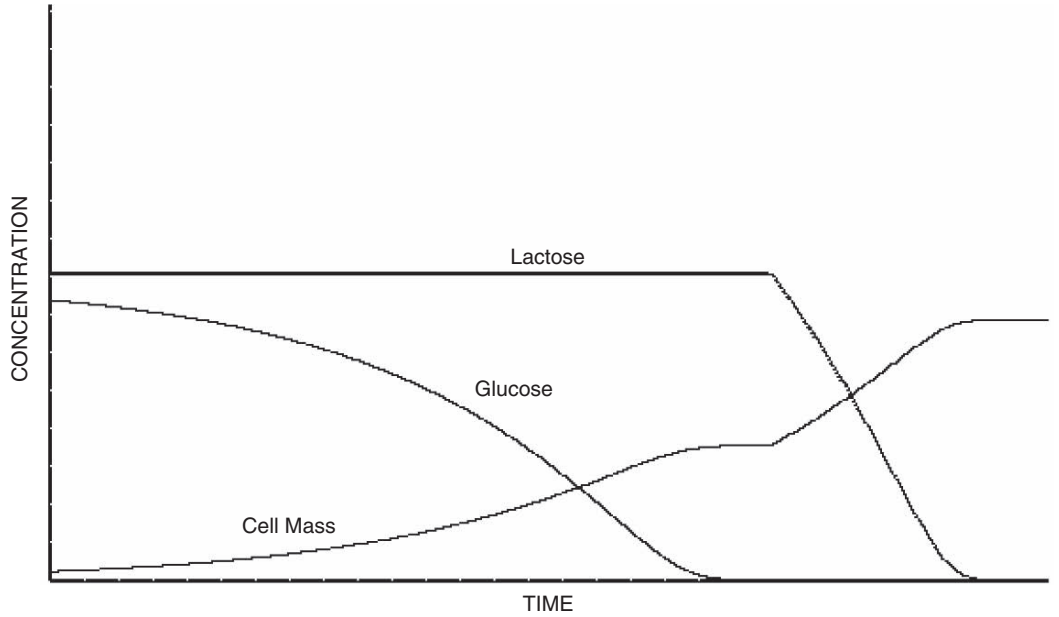

Fig. 1.20. Computer simulation of diauxie.

glucose and lactose, diauxie may not occur. Other ingredients such as glucose and glycerol are used simultaneously instead of sequentially because common organisms have the enzymes for converting both to intermediates in the main metabolic pathways.

When microorganisms suddenly encounter nutrients in excess concentrations for their metabolic rate, the cells absorb them for storage. Uptake by the cells is very rapid, and the process is called *biosorption*. This effect is used in a biological waste treatment process known as *contact stabilization* for rapid removal of organic materials. The high rate permits a short detention time, thus a small, inexpensive basin can be used. The cells are concentrated by sedimentation and sent to an aeration tank. A period of aeration promotes metabolism of the stored organic materials so that cells are ready for return to the contact basin.

Waste treatment media are relatively dilute; the low oxygen demand makes it fairly easy to maintain adequate oxygen concentrations. However, the delicate balance between supply and demand causes dissolved oxygen concentration to be an excellent, sensitive indicator of change of feed rate or feed concentration.

6.6. Biological Interactions

Very little is known about fundamental characteristics of mixtures of microorganisms; basic research is needed on growth rates, survival, population dynamics, physiology, and ecology. There would be little relevance to following the early ecologists in simply isolating and characterizing the constituents of a mixture. Although identification is essential, pure culture studies do not explain the mixed population. The aim is environmental control of the activity

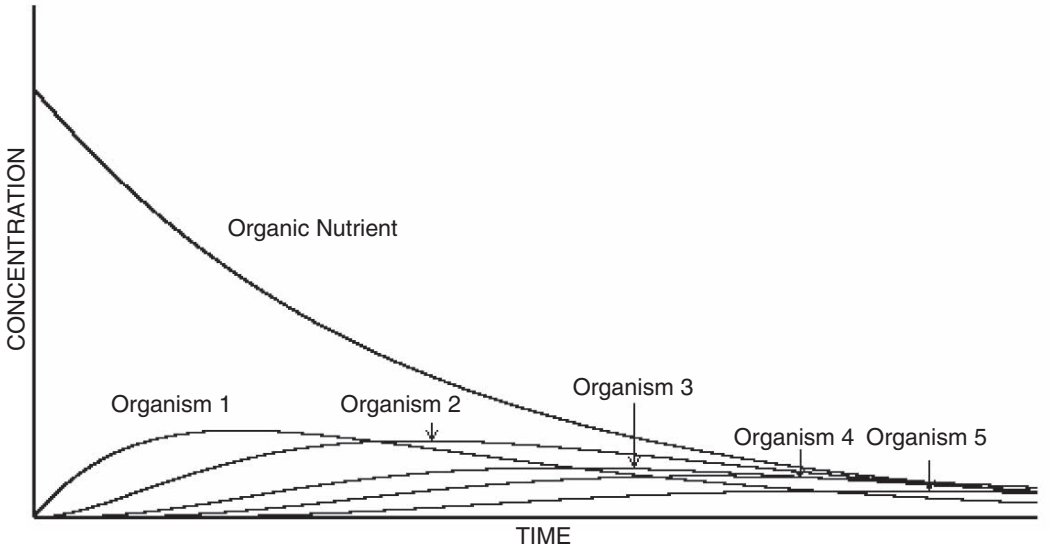

Fig. 1.21. Computer simulation of changing predominance of organisms in a batch treatment of organic wastes.

of populations rich in the desired organisms. To achieve this, it is necessary to know (1) the desired organisms, (2) rate or yield limitations, and (3) how energy is transferred.

6.7. Ecological Succession

Commonly, in natural environments one type of organism occurs in great abundance for a while, and then declines precipitously as another organism assumes predominance. A crude graph of possible population changes with time in a batch of nutrient material with a mixed inoculum is shown in Figure 1.21.

This figure is drawn by an internet exercise at: http://www.rpi.edu/dept/chem-eng/Biotech-Environ/MixCul/success.html. The equations for this simulation account for growth and for a dying organism becoming food for others.

Open, continuous flow systems can also exhibit succession, as shown in Figure 1.22.

One type of organism can be extremely numerous for a few days and then fall to concentrations that are below the detection limits for the analysis. Another species may dominate for a few days and then be replaced by still a third. The fundamental differences between batch and continuous systems are: (a) predominance in continuous culture depends on the outcome of many different associations instead of organisms dying to become the food for the next population; (b) a species often reappears in continuous culture after a period of being disfavored; and (c) the continuous-flow culture medium will exhibit a much narrower range of fluctuations than observed in a batch.

One aspect of ecological succession has very direct application to biological waste treatment. During startup of a process, batch operation is often practiced to build up a population

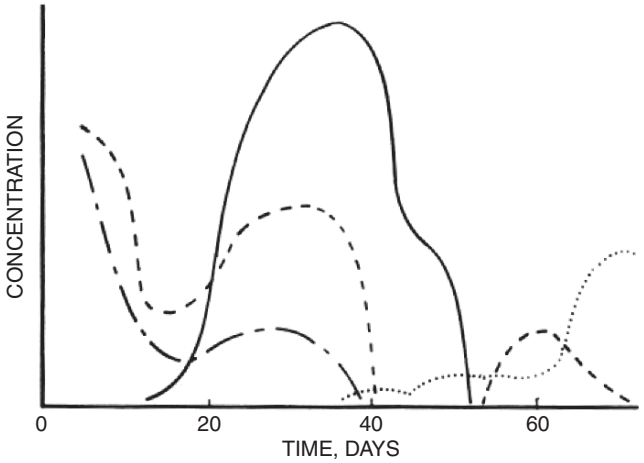

Fig. 1.22. Species variation in a continuous flow system.

with suitable characteristics for efficient metabolism of wastes. The problem is to initiate continuous feeding when the desired population is present. Unfortunately, the desired properties of a population are not well known, thus startup is usually a hit-or-miss proposition with uncertain results.

7. CONCLUSIONS

Biological waste treatment is complicated and poorly understood. At times, process performance can be unsatisfactory, but no explanation can be found. An attractive alternative to biological waste treatment is chemical coagulation followed by sedimentation of the precipitates, and such a process behaves quite reliably and can use automatic controls. Chemical treatment can be preferable when toxic compounds or poor nutritional balance prevent or retard microbial growth. However, chemical sludges, which are the solids from the wastes plus the chemicals for coagulation, tend to be voluminous with little potential value. Microbial sludges are being considered as energy resources and have established value for soil conditioning or as low grade fertilizer. As better understanding leads to more reliable performance, biological waste treatment processes should outperform chemical coagulation, except for some specialized industrial wastes. Nonbiological processes such as chlorination and carbon adsorption will, of course, remain important as supplements to biological treatment. The membrane operations of reverse osmosis and ultrafiltration are expected to play increasing roles, but could be used in conjunction with biological waste treatment.

Few fundamental improvements have been made in recent years in biological waste treatment. As bioengineers apply the principles of biochemistry and biology, and as mixed cultures become better understood, the future of biological waste treatment seems bright. For additional technical information on this subject, the readers are referred to the literature (1–20).

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2
Treatment by Application Onto Land

Donald B. Aulenbach and Nicholas L. Clesceri

CONTENTS

Abstract Various land application technologies as means for waste treatment and disposal are introduced. This chapter discusses the engineering topics of: surface spreading, overland flow, infiltration, percolation, slow rate irrigation, crop irrigation, vegetative cover, soil adsorption, soil ion exchange, mineralization, bacteriological activity, waste treatment, water removal, nitrogen removal, phosphorus removal, metal removal, disinfection, process design, land application rate, area requirements, monitoring, organic loading, siting, costs, etc.

Key Words Land application • waste treatment • waste disposal • surface spreading • overland disposal • slow rate irrigation • crop irrigation • vegetative cover • soil adsorption • rapid infiltration • percolation • siting • land requirements • nutrient removal • design.

1. INTRODUCTION

1.1. Scope

Various land application techniques for treatment and disposal of wastewater are presented. Considerations include the disposal of the water, removal of pollutants in the wastewater, the impact upon the soil through which the liquid is passed, and potential groundwater contamination. Means discussed for the disposal of wastewater are divided into (a) surface spreading or overland flow, (b) slow rate or crop irrigation, and (c) rapid infiltration-percolation. There are variations of these techniques, which are considered in terms of these three mechanisms.

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Application in wetlands and subsurface disposal are not considered here. The advantages and disadvantages of applying the liquid onto ground that has a plant cover as opposed to application onto bare ground are considered. No one method is recommended as preferred, because the optimum methods are usually site specific. All factors, including economics, must be considered before a final decision is made to use any method of land application of wastewaters.

1.2. Philosophy

The application of human and other animal wastes to soil as a nutrient or fertilizer is not new. History is replete with examples of the use of these wastes, from the Chinese who regularly collect "*night soil*" for direct use on their farms, to the present-day Amish people who have some of the most fertile soils in the nation owing to their "*organic farming*" using animal wastes. However, it was not until the 19th century, when the system of water carriage of wastes became popular, that the added problem of disposal of the liquid arose. In general, however, water carriage of wastes predominated in larger cities where the greater distance to the farmlands made it uneconomical to transport the total mixture back to the farms. Contrariwise, in rural areas where the domestic sewage would have been beneficial to the farms, there was generally insufficient sewage available. Thus, disposal of domestic sewage onto the soil did not receive wide usage.

Pound and Crites (1) have summarized the historical records on sewage farming. The earliest recorded use of sewage for farm irrigation was at Bunzlau, Germany in 1559. In the United States the first recorded use in irrigation was at Augusta, Maine in 1872. Many of the original sewage farm systems have been abandoned, primarily because of the growth of the cities they served. As the population expanded toward the farms, the value of the land for building became greater than the value for farming, and the farms were sold.

The disposal of wastes onto the soil in relatively modern times is not completely absent. At Lake George, NY, settled, unchlorinated trickling filter effluent has been discharged onto a natural delta sand deposit since 1939 (2–5). The system at Seabrook Farms, NJ, has been in use since 1950 (6). This involves the spray irrigation of food processing wastes back onto the land on which the crops grow, plus onto forested areas. Considerable study was done at the Flushing Meadows, AZ, rapid infiltration site to determine contaminant and nutrient removal or travel in a sandy soil (7). Other studied installations include the irrigation systems at Muskegon County, MI (8), Roswell, NM (9), Dickinson, ND (10), Tooele, UT (11), and Tallahassee, FL (12); the rapid infiltration systems at Vineland, NJ (13), Milton, WI (14), and Boulder, CO (15); and the overland flow system at Pauls Valley, OK (16).

With increasing concern for the quality of the environment, and particularly from the influence of Public Law 92-500 and Public Law 95-217, society has begun to look for new ways to improve its waste treatment and disposal methods with an eye toward reusable water recovery. Engineers have learned to cooperate with nature in constructing biological treatment systems that increase the efficiency of treatment. However, these were found not to produce the desired degree of treatment within the time and costs that were allowed, particularly for the soluble inorganic nutrients nitrogen and phosphorus. Chemical treatment is being used increasingly to improve the degree of treatment. This, however, has been found to be costly

in many instances. Therefore, other methods, such as the disposal of wastewater onto soil, are being investigated to achieve a greater degree of treatment at reasonable cost. Thus there has been increased popularity of land application of wastewaters.

Another factor influencing the consideration of soil treatment and disposal of domestic wastewater has been the ecological aspect. Wastewater contains material in the form of garbage, some partially digested food, used food or waste products, many minerals, and innumerable microorganisms. Much of the organic matter in the initial food was derived directly or indirectly from the soil in which it was grown. In land disposal of wastes, the soil is replenished of the nutrients that were removed by the growing plants (17), because the organic and inorganic constituents in wastewater are very frequently the same materials required for plant growth. Further, biological systems are available that can convert other materials in wastewater into usable nutrients for plants. Thus it seems to be appropriate to attempt to apply domestic wastewaters to the soil to complete the nutrient cycle. This alternative to incineration, disposal at sea, or ultimate disposal by other methods returns useful materials to the soil. It must be cautioned that soil disposal may not be applicable for the disposal of certain industrial wastes. Industrial wastes may contain heavy metals, toxic materials, and possibly such persistent or harmful substances as pesticides in sufficient quantity to cause injury to the soil system or a potential buildup of such things as heavy metals in the plants grown in that soil. These substances could be further concentrated in passing through several biological systems before reaching the human body. A solution to this problem is to use such wastes for irrigating trees and other plants that are not consumed as food.

Cost may be a limiting factor in the use of irrigation for wastewater disposal. The gregarious nature of humankind has prompted people to live in cities far from where their food crops are grown, thus creating difficulties in the transport of fertilizing materials back to the farm where they are needed. However, the important factor considered here is that irrigation is a valid alternative for the treatment and disposal of domestic wastewater and some industrial wastes.

One caution, however, must be taken. At present, our knowledge of all of the possible ecological effects is not complete. Therefore, we should not merely pursue irrigation as the ultimate means for disposal of all wastes. Much information is already available, but more must be secured and specific problems must be considered individually. Caution must be taken to prevent a buildup of harmful or toxic materials in the soil, or an upset of the salt balance in irrigated crops (18). These are effects that may not appear immediately and they must be evaluated over a longer period of time.

Sight must not be lost of the advantage of storing water in the soil. In many instances, particularly in rural areas, the major source of water supply is ground water that, after being used, is frequently discharged to the nearest surface water stream, with a concommitment lowering of the ground water level. Just as it is usually recommended practice to return surface water, after use and renovation, to the original watercourse from which it was secured, it is also advantageous to return ground water to the ground. There are many advantages to storing and moving water in the ground. These include the relative lack of evaporation, the cooler temperature, the relatively slow rate of flow, and the natural ability of the soil to filter and purify the water.

There are many advantages to the use of land application techniques for the disposal of the liquid effluent, for the purification of wastewater, and for soil fertilization (18–21). Some of the advantages and problems of this system are discussed in this chapter.

2. TYPES

2.1. Surface Spreading

Surface spreading or overland flow are the names applied to the technique in which the liquid is applied on the surface of the ground, with treatment being achieved as the liquid flows by gravity down a gentle slope. Application may be by a sprinkler system or by means of a pipe or a siphon at or near the surface of the ground, usually along an edge of a field, although it may be applied in a corner. Figure 2.1 (22) shows a sketch of a typical overland flow system. The liquid applied at the upper portion of the slope flows by gravity as a thin sheet or film to the lower lying sections of the field. The system is designed not to have any significant depth of water on the field at any one time or place; however, small indentations on the field may create minor puddles of water. By design, application rates are kept low enough to prevent overflows from the field. If there is an overflow, provisions are usually made to collect this overflow, channeling it to where it can be further treated or reapplied to another field (23). Application rates are a function of the slope, the type, and the character of the ground cover, the porosity of the soil, and the wetting caused by precipitation. In cold climates, freezing is usually a significant factor. Intermittent application is most desirable to allow time for the soil to lose much of its moisture before the next application. However, if bare ground is being used, the soil should not be permitted to dry entirely before the next application of spreading. The steepness of the slope of the system is limited by the point at which soil erosion takes place, and so is a function of the type of soil available. The method has its greatest application where gentle, even slopes are available and is generally more applicable where there is some type of vegetative cover on the ground.

2.2. Slow Rate

Slow rate application is usually used in systems designed for the irrigation of crops or forested areas. The main goal is to provide sufficient moisture for optimum plant growth and to provide at best a portion of the fertilizer requirement from the nutrients in the applied wastewater. As a consequence of this process, the impurities in the applied wastewater are converted in the soil and taken up by the plants, and the water is removed through evaporation, transpiration, or incorporation into the plants. Figure 2.2 (22) shows a diagrammatic sketch of some variations of slow rate systems for irrigation. In general the goal of slow rate systems is to provide sufficient water for the crop, but not so much that there is significant percolation to the groundwater. Average dose rates are in the order of 5 cm/wk (2 in./wk) usually applied in one or two doses per week at rates less than the infiltration rate of the soil in use to prevent surface runoff. Dosing may be by means of siphons with ridge and furrow distribution systems, or by some means of spraying. Natural precipitation must be considered as part of the total water being applied.

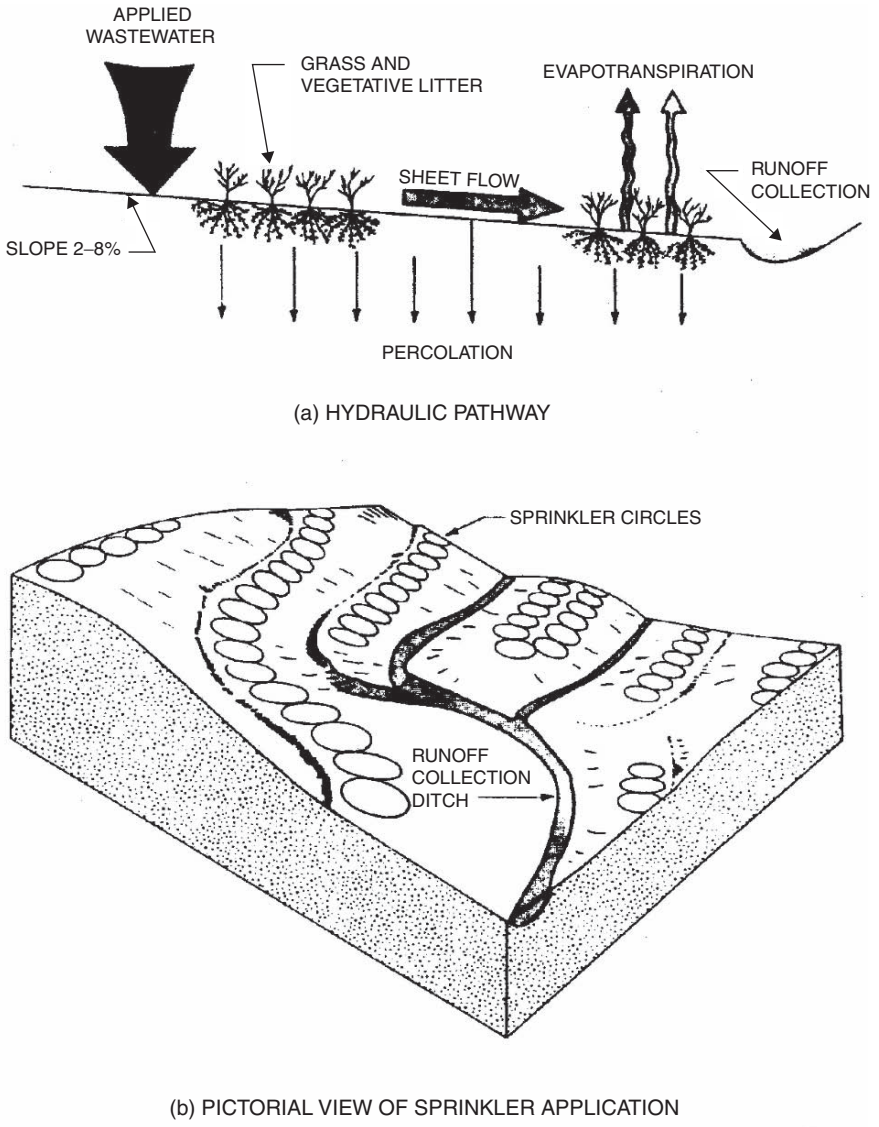

Fig. 2.1. Overland flow system (22).

Spraying includes all of the methods that involve discharging the liquid into the air and letting it fall to the ground, thereby resulting in some evaporation. Various techniques are used to achieve this effect. In general, moderately small droplets are preferred, although not a fine mist (24). Rotating nozzles, which give an intermittent effect, are preferable although fixed nozzles may be used. To obtain the desired spray, a piping system and a pump to develop the appropriate pressure are required. The most common system involves lightweight piping, frequently aluminum or plastic, with rotating spray nozzles at fixed locations. In some

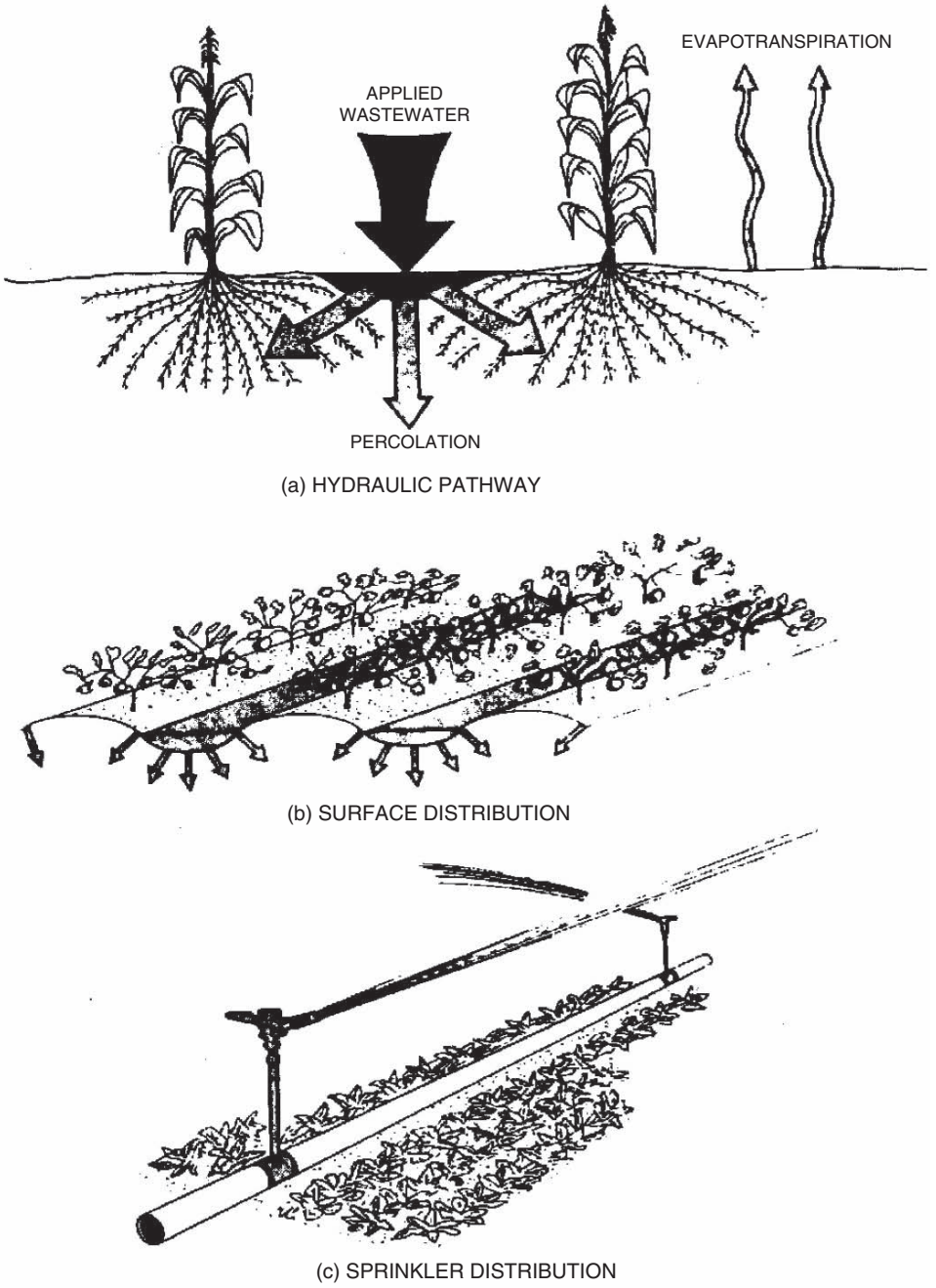

Fig. 2.2. Slow rate system (22).

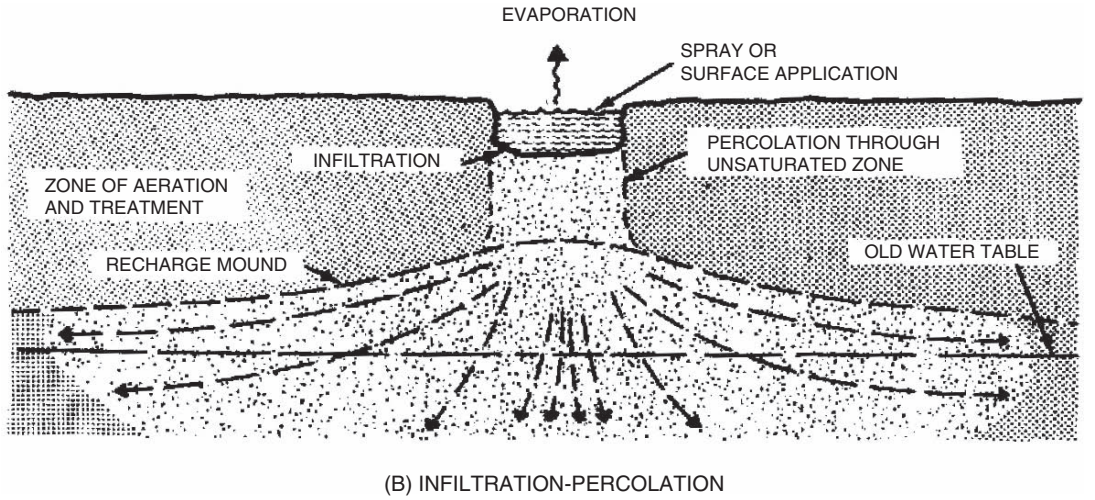

Fig. 2.3. Rapid infiltration-percolation system (26).

instances, portable piping that can be connected and disconnected readily for moving to another location is used. Flexible tubing, similar to a garden hose, has been used with the pressure of the spray causing the hose to be taken up on a spool, thereby moving the spray system. Other units involve long sections of elevated pipe that either rotate or reciprocate across a field. This latter system is not applicable where tall crops or trees are grown; however, spray irrigation of corn is very common. Regardless of the application method, reasonable caution must be taken to avoid the formation of aerosols that could be carried by the wind to an unwanted location (25). To help prevent this occurrence, spray nozzles are not set up at the immediate boundary of a spray site. Because pumping requires additional energy, the cost of a spray system is generally the highest of the systems described here, but spraying may be advantageous if a pump is needed anyway to move the liquid to the disposal site.

2.3. Rapid Infiltration—Percolation

Infiltration beds include all techniques wherein the water is applied to the soil to a measurable depth confined within embankments and allowed to infiltrate the soil and percolate to the groundwater (see Figure 2.3) (26). This usually implies a depression in the land to create the banks; however, artificially raised banks have been used. Infiltration beds, more commonly, do not have a vegetative cover. General design provides sufficient embankment to allow the bed to be filled from 15 cm (6 in.) to 100 cm (3 ft) deep with the applied water, which then seeps into the soil and thus into the ground water system (27, 28). An advantage of this system is that less continuous observation and manipulation are required, because a bed can be in a filling mode for a long period of time without concern that the depth of water may become too great. A complete system usually involves a series of beds filled alternately. This provides time for the applied liquid to enter the soil completely and allows the bed to dry and become aerated between dosing. This system is useful where there are significant changes in flow rate

over a period of time, but it is also susceptible to problems caused by precipitation. The beds are usually constructed by making depressions in the land. The most frequent reason for this is to remove the topsoil, which in general is a finer material and less pervious to the water than is the lower stratum. Where coarser materials may extend to reasonable depths from the surface, the depth of these beds may be fairly great. A specific use of this type of system is to divert storm runoff for gradual infiltration into the ground. There are more than 2000 such groundwater recharge basins on Long Island, NY (29). A modification is the ridge and furrow system, but the effectiveness of this system is essentially the same as for a normal infiltration bed. Ridge and furrow systems are also sometimes used for crop irrigation.

2.4. Vegetative Cover vs. Bare Ground

In all three systems discussed above, the surface may be covered with any vegetative system or the ground may be kept bare. There are specific advantages to and uses for each system.

The vegetative cover itself may be extremely varied. In some cases, the system may be designed to irrigate a specific crop. Different crops have different nutrient and water requirements; therefore, they have a varied capacity to renovate or purify the applied wastewater (30). In a total wastewater management system, it is important that the crops, with their incorporated nutrients, be harvested and removed from the irrigated field (1, 17). In other instances, the crop may be planted merely to increase infiltration of the applied liquid (31). Crops such as alfalfa have been found to increase the transfer of water into the soil. One of the disadvantages of annual plants, such as crops, is that they do not grow and take up the nutrients during the winter season. Thus where there is year-round discharge of wastewater, seasonal crops cannot be relied upon to take up and use the nutrients during the entire year. In this regard, the application of the liquid to a forested area has been found most acceptable, as described in the studies at Seabrook Farms (6). A forested area tends to keep the ground open better, even during freezing conditions (17). This is generally attributed to the penetrating root systems in the forest. Further, the nutrients are fixed in the soil both during the growing season and the nongrowing season. This is attributed to the litter on the forest floor, an organic-rich material that is capable of absorbing the available nutrients for the existing microbial systems that operate throughout the year, albeit at a slower rate during cold temperatures (30). Further, rotating nozzle spray systems have been able to be operated continuously in forested areas without the period of rest required for other systems.

Where uptake and removal of nitrogen are of concern, the use of a vegetative cover is most desirable (1, 20). This suggests that in summer resort areas, a vegetated system would be most amenable for the treatment of household wastes. Discharge of septic tank effluents should be at or very near to the ground surface so that the plants' roots can absorb the liquid and the nutrients, preventing their entry into the groundwater system. Obviously, this approach has some drawbacks where year-round living is conducted or where summer resorts are being converted to ski lodges.

Infiltration beds, on the other hand, are more amenable to a bare ground operation. Few vegetative species survive in the fluctuating flooded and dry conditions of an infiltration bed. Therefore, it would even be difficult to maintain a vegetative cover on such a system.

The question of vegetative cover is a function of availability and type of vegetation, the periodicity or time of year of discharge, the type and function of the irrigation system, and the type of soil available.

2.5. Final Residence of Liquid

There are two general possibilities for the final location of the applied liquid: evaporation-transpiration and groundwater recharge. In one the liquid never reaches the natural ground water, whereas in the other it is used to replenish it. The first is primarily concerned with preventing potential pollution of the groundwater aquifer relative to the type of waste being disposed into the soil system. The latter is frequently more concerned with replenishing the groundwater supply, but still with obvious concern for the quality of the ground water.

Evaporation-transpiration generally uses a vegetative cover and the application of only sufficient liquid to wet the root system of the plants being grown on the surface of the soil. Thus application rates are very dependent upon the vegetative cover and precipitation, and are generally in the order of 5 cm (2 in.)/wk. Intermittent dosing is used to allow the plants to use the water applied. Larger areas are required for this system, but greater purification of the waste and greater control of groundwater pollution are achieved.

Where the applied liquid is allowed to enter the ground water, the soil system's capacity to purify the potential pollutant is relied upon exclusively. This system is more concerned with replenishing the groundwater supply, but with obvious concern for the quality of the ground water. Greater control and inspection are required in methods where the wastewater reaches the natural ground water.

2.6. Chlorination

An unresolved problem is the necessity for chlorination of the waste before irrigation. Wherever there is the possibility of human contact with the wastewater, chlorination is highly recommended to control pathogenic organisms (32). However, much of the purification in soil is performed by biological activity that would be adversely affected by excessive chlorination. The greatest microbiological activity is near the surface of the soil, which is the portion that would be subject to the highest concentration of chlorine. Furthermore, chlorination adds dissolved solids to the wastewater and ultimately to the soil, and increases the cost of treatment. Some states require chlorination regardless of its necessity. In Michigan (33) chlorination is practiced for spray irrigation, but not for flood irrigation or seepage basins unless the effluent is recollected by underdrains. Thus, it may be seen that the specific conditions under which the irrigation is conducted will govern the necessity for chlorination.

3. PROCESSES

The most significant processes in the treatment of wastewater by land application are discussed here. The complete processes involved in the renovation of the water, the carrying off of the liquid, and the evapotranspiration of the liquid are not fully understood. Furthermore, there are numerous processes that are interrelated. In some instances, certain processes are

more responsible for specific removal abilities, thus the deciding factor as to which system is most desirable may depend upon the specific substance of concern.

3.1. Physical

There are numerous physical parameters that control purification by land application, with temperature being an all-important one (13). Temperature inversely affects the viscosity of the liquid being applied to the soil. The extreme condition, of course, occurs when ground is frozen, thus completely preventing liquid from entering it. Such conditions do exist in temperate climates during the winter and in colder climates for longer periods of time. In general, however, the temperature of both sewage and the water within the ground is relatively constant compared with the temperature variations at the surface of the ground. Temperature also affects other processes, particularly biological ones (34). Thus temperature alone could be considered separately; however, it is not a process in itself, but merely affects other processes, and, therefore, is not considered separately here.

3.1.1. Infiltration

3.1.1.1. Soil Type

The factor exerting the greatest control over infiltration is the type of soil (23), including the size of particles, where the optimum is somewhat in the middle of the range with poorer conditions at both ends of the particle-size scale (1). Clay has the smallest particles. These particles are of such size that the void space between them is small; as a result the water molecule itself finds difficulty in passing through. Although a clay soil can hold a large amount of water, the most important factor here is the transmissibility or the movement of the water through the soil. As the size of the soil particles approaches that of fine sand, the infiltration rate increases. Coarse sand allows even more rapid infiltration and small gravel approaches the optimum rate. As the size of the gravel approaches that of stone and rocks, the infiltration rate is decreased again because there are fewer pores or spaces between the rocks. Ultimately, a solid rock material size is approached, which restricts the flow completely because, generally speaking, rock is impervious to the passage of water. A potential exception to this occurs where rock has become fissured after any of several geological occurrences, including earthquakes, subsidence, and so on. Of particular note in this regard is limestone, which is relatively soluble in water and tends to become highly fissured by the passage of water through it. Large channels may appear in limestone, allowing a rapid flow of water through them with minimal contact with any surface. When considering only the removal of water, a large fissured rock may provide the best means of accomplishing this. However, in general, water renovation is also of concern and this type of formation provides the least treatment of the wastewater.

In addition to considering the soil types, the depth of any particular type of soil is an important factor (17, 23). Very frequently, a site is chosen with coarse sand at the surface only to find that this extends to only a shallow depth and is underlain by a less pervious soil. For maximum liquid disposal, this is less than desirable. However, from the standpoint of uptake of water by plants near the surface, this provides an optimum situation. Frequently, lenses of soils of different permeabilities exist in a field. This creates problems in infiltration,

often resulting in ponding in the areas underlain by less pervious lenses. The depth of soil to the ground water table is also important, especially in wastewater purification (28).

Although somewhat related to soil size, organic matter within the soil is also a factor in infiltration. In general, organic matter is finely divided and comes under the category of small-size particles. However, organic matter may have a gelatinous or a mesh-type structure that results in a slower passage of the water through the soil. Furthermore, consideration must be given to the buildup of organic matter owing to the nutrients in the applied wastewater. Ultimately, there should be a balance between the buildup and the degradation of organic matter in the soil. However, there is an optimum range for this balance and conditions may be adjusted to either increase or decrease the organic matter loading by judicious application of the wastewater.

3.1.1.2. MOISTURE

Soil moisture controls infiltration. Consideration must be taken of precipitation as it affects soil moisture. The amount of precipitation must be included in the calculation of application rates for irrigation. In areas where considerable rainfall occurs over a long period of time, some alternate means of storage of the wastewater must be provided until the ground dries sufficiently for resumption of the application to the soil. The speed at which rainfall infiltrates into the soil is quite similar to the infiltration rate of the applied wastewater effluent. In general, for rapid infiltration systems, the large volume (depth) of wastewater applied makes the additional input from precipitation insignificant.

There is an optimum soil moisture for maximum infiltration that is related primarily to the soil particle size. When fine clay soil or soil containing clay dries out completely, it forms a dry cake that is difficult to remoisten, thus making it difficult to reestablish infiltration until the slow wetting process has been accomplished. If normal application rates are applied to such caked soils, this frequently results in excessive runoff during the beginning of the irrigation period. Thus for fine clay soil, or soil containing clay, it is desirable to resume application of the irrigation water before the soil dries out completely.

3.1.2. Filtration

Filtration implies physical particle removal. It is obvious that the size of the particle to be removed from a wastewater is an important factor in considering this process. Very large particles will be trapped on the soil surface and will cause no significant problems, but may cause channelization of the applied effluent away from that particular area, thus reducing the total effective infiltration area and resulting in higher dose rates in adjacent areas. Generally of greater concern are the more finely divided particles. Here the size of the pore space between the soil particles is the most important factor governing the removal of particles. As a general rule, the size of the particles trapped is somewhat smaller than the pore size of the soil. This is caused by other factors, including adsorption and the presence of gelatinous biological masses onto which the particles will adhere, thus separating them from the liquid portion.

In general, if organic material is applied to a soil, a mass of gelatinous biological material accumulates on the surface. This in itself tends to filter out more of the finer particles, thus improving the degree of particulate removal and the clarity of the water that passes through

this system. On the other hand, this surface slime tends to clog the soil, thereby restricting the flow through it. Intermittent application to such a system allows this biological mat to dry out and the surface to be rejuvenated for another cycle of liquid application (35). When infiltration beds are constantly filled, algae tend to proliferate in the overlying water. This is filtered on the surface of the bed as the water infiltrates into the soil, thus clogging the bed. In some instances, raking and disking are required to reopen the pores of the soil. In extreme instances, a certain amount of the surface of the bed must be removed from the system along with the clogging mat to rejuvenate the infiltration capacity (3).

3.1.3. Adsorption

For adsorption, the consideration of soil type must include both the size of the soil particles and their chemical composition. Size is an important factor, inasmuch as adsorption is a surface phenomenon, and the finer a particle is divided, the greater the surface-to-volume ratio (36). Specifically, the process of adsorption relates only to sorption on a surface, but in a soil system such as is being considered here, both adsorption and absorption occur, and no differentiation is made here between the two processes.

Another factor in adsorption is the chemical composition of the soil particles. In general, the higher the valence of the elements within the soil, the greater is the adsorptive capacity. Thus, aluminum-containing soils have a tendency to display greater adsorptive capacity than predominantly sodium- and potassium-containing soils (36). Iron- and calcium-containing soils also possess significant adsorbing powers. Certain elements have more specific adsorption capacity and therefore the degree of adsorption on the soil is also a function of the ions adsorbed from the applied wastewater. Thus a generalization cannot readily be made for the adsorptive capacity of a soil without first making a thorough study of the materials that could potentially be adsorbed. Space here does not permit a complete listing of specific contaminants that would be adsorbed by specific elements within the soil; however, these combinations may be found in the literature (37).

Biological adsorption may also be significant in a soil. This is particularly true where nutrients, both inorganic and organic, are available in the applied wastewater, thus providing sufficient growth material for microbes (principally bacteria) in the soil. Bacteria themselves are very small particles and therefore have a relatively large total surface area on which adsorption can take place. For biological systems, adsorption is the first and most important step in concentrating the required substrate (including nutrients) at the surface of the organisms so the nutrients may be absorbed for growth and development. This includes larger biological systems such as roots, which rely upon root hairs for the increased surface area onto which adsorption can take place.

3.2. Physical-Chemical

3.2.1. Ion Exchange

In general, ion exchange involves adsorption of the elements onto the surface of the exchange medium followed by exchange of ions between the liquid and solid portion of the system (36, 38). Characteristically, the higher the valence of the element, the greater will be its force of attraction to the solid phase (39). Thus a trivalent ion such as aluminum would tend to

be removed from the soluble phase onto the solid phase and exchanged for three monovalent ions such as sodium, which would go into solution (39). Divalent ions such as calcium and magnesium would have slightly less force of attraction to the solid phase, and there are even certain monovalent ions that may be exchanged with other monovalent ions from the solid portion. There is a balance between the specific ions on the soil particles and those in solution (40). Many naturally occurring ion exchange sands exist, with a predominance having cation exchange capacity (41). As a matter of fact, the first ion exchange media used were naturally occurring sands. However, in a soil system such as that considered here, there is little or no opportunity to regenerate the ion exchange capacity, nor is it particularly desirable. As the ion exchange capacity is exhausted, higher valence ions will be less removed and therefore will be transported farther into the soil system. Unless some other process interferes, this will ultimately result in the exhaustion of the ion exchange capacity of the soil with a concurrent lack of removal of the specific substance.

There is some minor opportunity for regeneration of the ion exchange capacity of soil (36) when rainwater falls upon the area. The rainwater is lower in total ionic strength and therefore the ions will desorb from the solid phase back into the liquid phase to maintain equilibrium. Where high rate infiltration is practiced, the amount of rainfall compared to the amount of sewage applied is generally very small (42). However, with the application of relatively small quantities of wastewater, such as in crop irrigation, the amount of desorption and regeneration caused by rainfall may be a significant factor.

3.3. Chemical

There are many chemical processes going on within the soil system; however, of all these processes, mineralization is the most significant and therefore is the one discussed here.

3.3.1. Mineralization

Mineralization involves the desorption of certain adsorbed and ion-exchanged ions from the soil particles back into the liquid phase where they may combine with another ion to form an insoluble precipitate. This results in renewing the adsorption and ion exchange sites, thereby allowing a continuous process to exist (36). It is thought that this process is responsible for the continued capacity of a soil to remove phosphorus. Details of this are discussed in Section 3.5.3.

Mineralization can also refer to the process of conversion of organic matter to inorganic matter. This goes on within the soil, particularly when wastewater containing organic matter is applied to the soil. This mineralization usually involves biological processes and it is difficult to separate chemical and biological processes of mineralization in natural environments.

Another form of mineralization is the direct chemical combination of various substances within the soil-water system (36). Some of these combinations may be slow reactions that take place during the process of renovating the wastewater. This results in the formation of insoluble compounds from the combination of the soluble substances in the applied wastewater and the soil materials that become dissolved in the liquid. In these processes pH, buffer capacity, and redox potential play important roles (36, 38). In most wastewaters, the carbonate-bicarbonate buffer system is the most predominant and the species distribution is affected

by small changes in pH. Application of lime to the soil as part of the normal fertilization procedure may affect the pH significantly. A small change in pH could liberate certain ions, specifically calcium, which would then be available to react with other ions, phosphorus, as an example, forming an insoluble calcium phosphate under satisfactory pH conditions. The solubility of iron and manganese is controlled by the redox potential (37). Many other reactions may also take place, resulting in the solubilization or precipitation of substances in the soil-water system.

Thus it may be seen that mineralization plays an important part in the overall improvement of the quality of wastewater applied to the soil.

3.4. Biological

Many biological systems exist in soils. These range from viruses through the bacteria, fungi, algae, and finally to vascular plants, including crops and trees. Each system has a unique biological process and many of the processes are interrelated. With regard to the higher forms in particular, there can be a great range of both infiltration and purification accomplished by specific crop cover, and even greater differences when trees are used as a ground cover. Here again, space prohibits the discussion of every biological form. However, the most significant factors, the vascular plants and the bacteria, are discussed in some detail.

3.4.1. Plant Life

3.4.1.1. TREES

Trees play a very important part in the success of a land application system. From the standpoint of infiltration, the roots tend to keep the ground open and allow for rapid entry of the wastewater into the soil, even during freezing weather.

For the ultimate removal of water, trees also bring into play their great capacity for transpiration. Large amounts of water are taken up during the growing season and passed from the root system through the entire tree to the leaves. In general, deciduous (leaf-bearing) trees do not transpire much water during freezing conditions; however, many coniferous ("evergreen") trees do continue transpiring even during severe freezing conditions.

Another function performed by trees is the uptake of nutrients for plant growth. This may include both organic and inorganic materials. In trees, this is converted to two types of growth. One is the more permanent woody material composing the trunk and the branch system. The other is the leaf system that is dropped back to the soil. This leaf litter provides a thick mat that retains nutrients, encourages biological growth, and maintains the soil pores in an open condition. As a result of this mat, spray irrigation in wooded areas may continue throughout the winter, permitting continuous operation of the disposal system.

Trees exhibit a wide range in uptake of both nutrients and moisture (43). Where irrigation of existing woodland is to be practiced, there is little choice of the type of trees used. However, in establishing new infiltration areas, specific trees may be planted to provide the forested area. In such instances, consideration should be given to trees that are particularly suitable for the climate, the soil type, and the amount of moisture to be applied. Studies have been made of sewage purification by trees from the cypress of the tropical swamps (44) to the pine trees of the cold northland. In general, the application of the wastewater tends to increase the rate

of growth of the trees. In addition, where sandy soils are encountered, improvement of the soil quality has also been observed. Thus irrigation of forested land has been shown to result not only in increased purification of the wastewater applied, but also in improvements in both the soil and the forest crop (43).

3.4.1.2. Crops

A wide variety of crops may be grown on land irrigated with wastewater effluent. Consideration must be given to the quality of the wastewater applied and the use of the crop. In general, it is recommended that wastewater be chlorinated before being applied to crops that are used for direct consumption. Furthermore, a crop consumed without cooking, such as lettuce, tomatoes, and so on, should not be irrigated with wastewater effluent for a period of 1 to 2 weeks before harvesting, depending upon the soil moisture conditions and the precipitation at the time (25). A crop that is to be further processed before consumption, particularly by heat treatment or cooking, or one that passes through other animals before human consumption, may be irrigated with wastewater up until the time it is ready for harvesting (25). Crops such as grain and hay require the termination of irrigation before the harvesting time to allow the soil and the crop to dry before harvesting. Irrigation should not be resumed until a crop such as hay has thoroughly dried and has been removed from the field. Irrigation also should not be applied immediately before plowing and seeding a field because of problems with the use of machinery in a wet field.

Special situations may arise, such as those observed at the Seabrook Farms. Here the wastewater to be used for irrigation came from the processing of earlier crops harvested from adjacent fields. After the first crop, processing wastewater for irrigation was generally available for later crops. Thus there must be some temporal coordination of irrigation with the crops grown.

In general, for the irrigation of crops, a lower application rate, with intermittent periods of drying, is more advantageous. Complete drying of the soil is not recommended and the irrigation must be coordinated with normal rainfall. Just as the uptake of nutrients and water among various species of trees is varied (43), so too is the uptake by various crops (1, 43). However, more important with crops is the ability of the plant stems and roots to keep the soil open so that the irrigation water may infiltrate the soil.

Irrigation of crops is more amenable to a system of limited irrigation. This is a system whereby approximately 5 cm (2 in.)/wk of irrigation water is applied to the crop in an effort to restrict the irrigation water to only the root zone of the crops. The roots will then take up both the water and the nutrients, converting them into plant growth. Particularly where there is concern for the continued use of the soil for crop production, care must be taken not to apply an excess of irrigation water containing significant amounts of dissolved solids. The plants may not take up all the dissolved solids, thus resulting in their accumulation in the soil. If more solids accumulate in the soil than are taken up or leached out by means such as flooding or precipitation, there may be a buildup of salts in the soil that will render it unusable for future crop production (30, 31, 38, 43, 45, 46). Specific crops have varied uptakes of dissolved salts and it is frequently possible to find another crop or strain of a specific crop that can tolerate the increase in salt concentration in the soil (1). However, there are limits to

the crops' tolerance and uptake of salts, and precautions must be taken not to exceed these limits.

Consideration must be given to specific goals in the use of wastewater in irrigation, such as renovation of the wastewater, removal of the water, an increase in crop production, or the improvement in soil quality. Each condition will have a different optimum rate and method of application of the wastewater.

3.4.2. Bacteriological Activity

Soil contains many bacteria, fungi, and other microorganisms. In addition, there are enzymes related to these bacteriological systems. This section will discuss only the action of bacteria because they play a prominent role.

In general, the greatest numbers of bacteria in soil are concentrated near the surface (23, 25, 34, 47). The soil tends to act as a filter for bacteria and therefore, relatively few bacteria applied at the surface of the soil, either by natural causes or by irrigation, penetrate deep into the soil. However, this is also a function of the pore space within the soil. Coarser soils have larger pore spaces and will permit bacteria to penetrate to greater depths (34). A factor controlling the numbers of bacteria is the availability of organic matter and mineral nutrients (34). Where forest cover is present, the leaf litter on the forest floor provides a thick mat of organic matter in which the bacteria can concentrate. In order for heterotrophic bacteria to survive within the soil, they must have a source of organic matter that provides their energy requirements. Though there are a few autotrophic bacteria present that can survive on inorganic material alone, the energy conversion of this component is small and relatively large amounts of inorganic materials must be made available. However, with high irrigation rates, small concentrations of organic matter and nutrients may provide sufficient amounts to support both types of bacterial growth.

The bacterial system includes the associated enzymes. These enzymes are capable of breaking down larger organic molecules into less complex ones that can be used as food by the bacteria. Cell growth of bacteria entails the synthesis of large organic molecules. However, as the available organic matter is depleted, the bacteria die and themselves become organic matter for other bacteria. Meanwhile, each step along the way results in the ultimate conversion of some organic matter to mineral matter. This mineral matter, in turn, can be used as nutrients for the growth of green (photosynthetic) plants. In general, this involves the trees and crops as discussed previously, but can also include the growth of algae. Algal growth within the soil is limited to regions very near the surface because of the obstruction of light by the soil particles. If the drainage or runoff from an irrigated field collects in a nearby stream or lake, the nutrients can stimulate algal growth there. The net effect of bacterial assimilation and transformation of organic material is to act as a system to provide inorganic material for use by other biological systems (48).

There are many bacteriological systems at work in soil. However, in general, they may be classified into three main categories, according to their oxygen requirements: aerobic, requiring oxygen, anaerobic, requiring the absence of oxygen and. facultative bacteria falling into either category. In general, the aerobic bacteria metabolize organic matter much more efficiently and convert more of the organic matter to mineral material. Furthermore, they

produce stable minerals that are less objectionable and harmful than the products produced by anaerobic systems. The products of microbial metabolism of organic compounds in soil under aerobic and anaerobic conditions may be shown by the following generalized equations (47):

Aerobic (1)

$$(CHO)_nNS + O_2 \rightarrow 60\%\ CO_2 + H_2O + 40\%\ \text{microbial cells} + \text{storage products} + NH_4^+ + H_2S + \text{energy}$$

$$\downarrow \qquad \downarrow$$

$$NO_3^- + SO_4^{2-}$$

Anaerobic (2)

$$(CHO)_nNS - O_2 \rightarrow 20\%\ CO_2 + H_2O + 5\%\ \text{microbial cells} + \text{storage products} + 70\%\ \text{organic intermediates} + 5\%\ CH_4 + H_2 + NH_4{}^+ + H_2S + \text{energy}$$

In addition, there is concern about pathogenic bacteria that may be present in the irrigation water. Pathogenic organisms, being accustomed to the anaerobic intestinal tract of warm-blooded animals, are normally of limited infectivity when they are subjected to aerobic conditions (25, 47, 49). Therefore, from the health standpoint, there is a great advantage to maintaining the soil in an aerobic condition, which enhances the removal of pathogenic organisms (18, 31). Besides encouraging the growth or persistence of pathogenic organisms, anaerobic conditions frequently result in the production of undesirable gases, such as hydrogen sulfide, and other materials that are obnoxious from an esthetic standpoint. Thus there is sufficient reason to try to maintain the soil in an aerobic condition during application of the wastewater (34). It is an over generalized feeling that soils unsaturated with water are aerobic, whereas soils saturated with water are anaerobic (46). This is not necessarily so, because oxygen consumption is frequently controlled by the amounts of organic matter and bacteria present (48). By proper application, some dissolved oxygen may be maintained in a saturated soil. Aerating the applied water will increase the dissolved oxygen transferred to the ground water. Spray systems provide significant aeration, and rotary sprays also provide intermittent dosing. Aeration within the soil may be achieved for both surface spreading and infiltration bed techniques by intermittent dosing. During the period of non-irrigation, the soil dries out and air may pass through the void space of the soil, completely aerating the soil and any water in it. Immediately upon reinitiating the irrigation, the air entrained within the soil is carried through the soil along with the water, thus maintaining aerobic conditions (31). Optimum conditions for the period of intermittency must be determined for each soil type, cover crop, and system of application of the wastewater.

3.5. Process Applications

Although space will not permit discussion of every specific process occurring within the soil, some of the more important and well-known processes are considered here.

3.5.1. Water Removal

One of the most important reasons for the use of irrigation systems for disposal of wastewater is to dispose of the water itself. There are three basic means of ultimate removal of the water: (a) evaporation, (b) transpiration, and (c) percolation of the water to the ground water. The first two methods return the water to the atmosphere, whereas the third carries the water into the ground water, which may or may not remain in the ground, depending upon the aquifer. There may be advantages and disadvantages to both of these repositories. Discharge into the atmosphere will result in increasing humidity and ultimately in increased precipitation downwind from the disposal site. On the other hand, there is great concern for the depletion of the groundwater supply; thus, returning water to ground water may be very desirable under certain circumstances.

Surface spreading is generally designed to provide water only to the depth of plant roots. Here, evaporation may take place through capillary action, carrying the water to the surface of the soil where it is evaporated. In turn, the roots will take up much water and transpiration will occur during photosynthesis. A small amount of moisture will be incorporated into the crop and this will be ultimately removed with the crop.

Spraying may also result in a large amount of evaporation. By spraying the liquid into the air, some evaporation takes place before the water reaches the ground. Some caution must be taken, because such evaporation also results in the concentration of the remaining dissolved materials in the wastewater. Also, there is concern about the formation of aerosols. The evaporation of the droplets results in the production of dust and other aerosol particles that may contain bacteria and/or viruses (25). This could be a potential means for transmission of diseases. If this is a serious concern, chlorination should be practiced before the spraying. Whether or not the sprayed liquid ultimately reaches the ground water is a function of the application rate. Spraying can be regulated either to prevent the water from reaching the ground water or to allow it to do so to replenish groundwater supplies.

Infiltration beds generally are designed to recharge ground water. In this system, the water is applied to the surface of the bed to a depth of usually less than about 1 m (3 ft) and the water allowed to infiltrate into the soil, percolating down to the ground water, and then mixing with and becoming ground water itself. Depending upon the porosity of the soil and the slope of the groundwater table, factors that control the transmissibility of applied water, this water may be carried great distances through the ground and used as a water supply a considerable distance away. Thus it may be seen that land application must be included in total water management.

3.5.2. Nitrogen Removal, Conversion, and Transmittal

Nitrogen may occur in many forms including organic nitrogen (RNH_2), ammonia (NH_3) or ammonium ion (NH_4^+), nitrite (NO_2^-), and nitrate (NO_3^-) (41, 50). The form in which nitrogen will appear is a function of its form in the applied wastewater, the redox potential, the availability of oxygen, and the bacterial system present (41, 51). Steps in the oxidation process are (36):

$$RNH_2 \rightarrow NH_4^+ \rightarrow NO_2^- \rightarrow NO_3^- \tag{3}$$

Organic and ammonia nitrogen are normal components of sewage, being products of animal metabolism and part of the protein found in food. Under aerobic conditions and in the presence of sufficient nitrifying bacteria, these may be converted to nitrite and ultimately nitrate (2, 31, 51). Although this is considered a desirable process, nitrates are completely soluble and may be carried great distances within the soil (1, 27, 41, 52–54). Furthermore, the presence of excess nitrates in water supplies has been linked to the disease methemoglobinemia, or the condition called "blue babies" (55). To assure that water is safe for infants, the US Public Health Service has set a limit of 10 mg/L of nitrogen as the limit of nitrate in drinking water (56). The buildup of nitrates during the passage of sewage or septic tank effluent through the soil has been observed (43, 57), requiring that nitrates be monitored to warn of potential contamination of drinking water supplies. The application of irrigation water to shallow depths where plant roots may use the nitrate as a nutrient becomes important in nitrate control (13, 17). Plants require relatively large amounts of nitrogen as a nutrient, and, in general, the amount supplied by a treated effluent is less than that required for optimum plant growth (20). Grasses have been shown to possess poor nitrate uptake (58). In the laboratory, 5 to 6 hours contact time was needed to achieve 90% nitrate removal. In other studies (43), reed canary grass showed a far greater uptake of nitrate than did corn. A problem arises in temperate climates, where the plants do not grow during the winter period. A solution is the irrigation of forested areas. The organic mat of the forest floor appears to be capable of taking up the nitrogen and storing it until active biological growth again occurs in the spring (30). Certain cover crop residues may also have a limited capacity for storage of nitrogen, but it is much greater in the forest floor cover. In trees, the nitrogen is stored in the woody material, whereas in crops the stored nitrogen is removed from the system when the crop is harvested.

Nitrates may be reduced to nitrogen gas under certain anaerobic conditions (31, 50), particularly where large amounts of decomposing organic matter are present, as in forest litter (1, 43). The reaction proceeds as follows (51):

$$NO_3^- \rightarrow NO_2^- \rightarrow N_2 \uparrow \tag{4}$$

This is considered a means for the transfer of nitrogen from the aqueous system to the atmospheric system. Whereas this may be effective in a bare ground system, it may not be effective where a plant cover is employed. Certain leguminous plants are capable of fixing atmospheric nitrogen into their cellular material, thus increasing the total amount of nitrogen within the soil system, according to the reaction (51):

$$N_2 \rightarrow NH^{2-} \rightarrow HN_2^- \rightarrow NH_3^- \rightarrow RNH_2 \tag{5}$$

Because the total nitrogen available in the irrigation water is frequently less than the requirement for optimum crop growth, when leguminous crops are grown, there may be an increase in the total nitrogen in the soil system (52). However, the net result is still a reduction in nitrogen in the aqueous phase.

Some of the nitrogen pathways and interrelationships are depicted in Figure 2.4 (55). In general, where removal of nitrogen is of concern, this is best achieved by applying the irrigation at a low rate to a system employing plant cover (17, 30, 59).

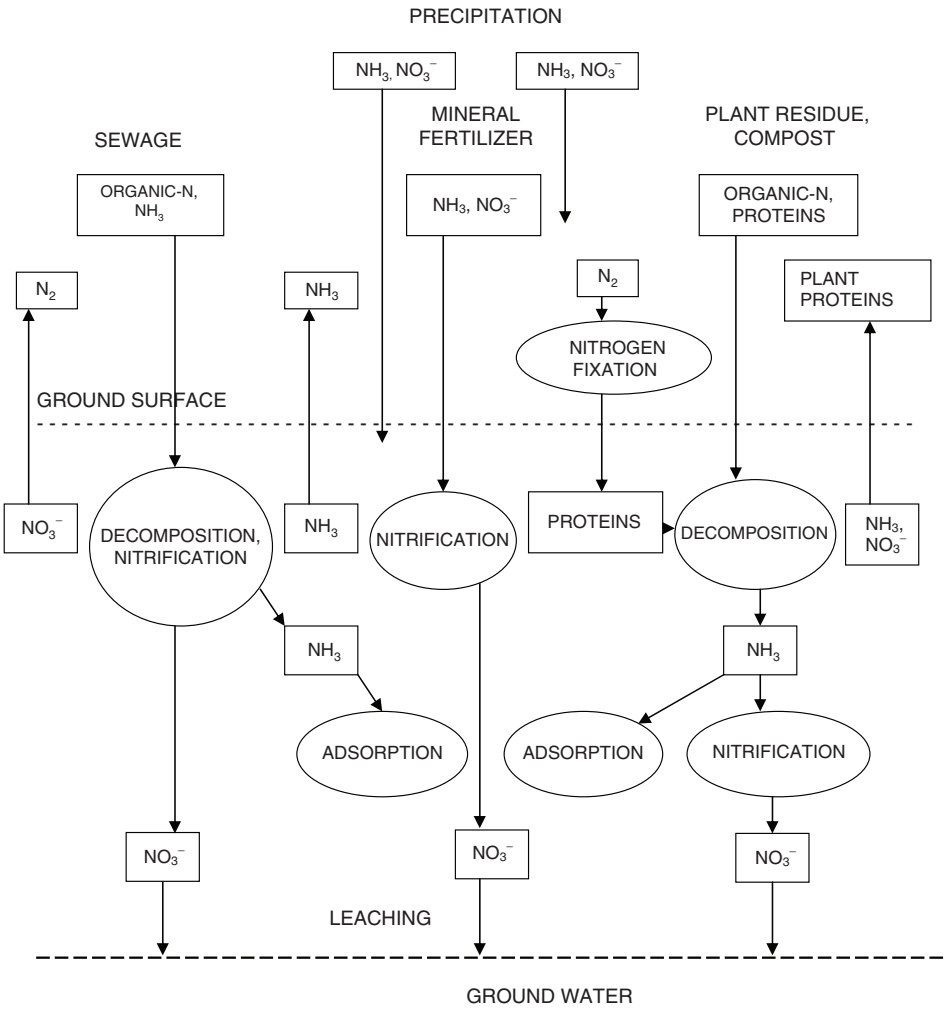

Fig. 2.4. Nitrogen transformations in soil (55).

3.5.3. Phosphorus Removal

There are two general processes by which phosphorus is removed from wastewater applied to the soil. The first is the uptake by any plants whose roots are fed by the applied wastewater (17, 20, 28, 42, 43). Phosphorus is an important nutrient for plants, so they will remove it from the water or the soil to supply their own energy and plant growth requirements. The other process is the removal of the phosphorus by the soil itself (1, 31). This is a much more complicated process.

Phosphorus removal in the soil appears to be a three-step process (28, 36), all of which may be going on simultaneously within any soil system. The first process is one of adsorption,

in which the phosphorus is removed from the liquid and is attracted to the surface of the soil particles. The size of the soil particles is a controlling factor because smaller particles have a greater surface area per unit of volume than do larger coarser soil materials. Certain clay materials, particularly montmorillonite and kaolinite, have a much greater adsorptive capacity than other clays of similar surface area (28, 36, 42, 43). The second process is ion exchange. After the phosphorus is drawn to the surface of the soil particle, an ion exchange reaction takes place in which the phosphorus, as phosphate having a valence of -3, is selectively exchanged for ions having a lower valence that are released from the soil particles back into solution. Thus the phosphorus is more firmly attached to the soil particle than merely by adsorptive forces. Studies have shown (60) that under natural conditions, even sand grains may have a coating of aluminum, iron, and calcium. Finally, the phosphorus is rendered inactive by a mineralization process. This may take two different routes. The more readily understood route is the simple precipitation of the phosphate with other ions in the irrigation water, specifically iron, aluminum, and calcium (17, 41, 47). Calcium forms the most stable combinations with phosphorus [$Ca_3(PO_4)_2$ and $Ca_{10}(OH)_2(PO_4)_6$], but its formation at or near neutral pH is far from 100% complete (61). Probably, the most common precipitate of phosphorus is an aluminum phosphate ($AlPO_4$). This is relatively stable at most pH ranges encountered within the soil and is also stable under both aerobic and anaerobic conditions (61). Iron phosphate ($FePO_4$) is also quite stable at extreme pH ranges (61), but under anaerobic conditions, the iron is generally reduced to the ferrous ion, in which case it becomes soluble, liberating any precipitated phosphate (28, 45).

The other more complicated process of phosphorus removal involves a form of reverse ion exchange, releasing the phosphorus from the soil particles onto which it had been exchanged. The exact conditions of this process have not been completely explained at the present time. However, its occurrence has been confirmed because soils whose ion exchange capacity has been measured carefully have been shown to continue to be able to remove phosphorus by ion exchange mechanisms many years after their ion exchange capacity would have been exhausted based upon the quantity of phosphorus applied to the soil (28). The process probably involves an equilibrium in which there is always a certain amount of phosphorus remaining in solution that can then be precipitated out as mentioned previously. This establishes a new equilibrium with new, although small, amounts of phosphorus. This gradually results in the total precipitation of phosphorus in the soil and the continued ion exchange capacity of the soil particles for extensive periods of time (45, 52).

Phosphorus, as nitrogen, is also an essential plant nutrient. Crops and plants will take up phosphorus, incorporating it into the cell material. With trees, this may accumulate in the woody material whereas with crops, significant amounts may be removed in the crop itself or with the silage that is removed from the field (62). However, if the crop is not removed, but is allowed to decay in the field, just as with fallen leaves from trees, the phosphate may be leached out of the vegetative portion back into the liquid portion (30, 63).

Thus, phosphorus may be removed effectively on both bare and plant covered fields (2, 20, 27, 42, 43, 52, 53, 57, 59). Some phosphorus interactions in soil are depicted in Figure 2.5.

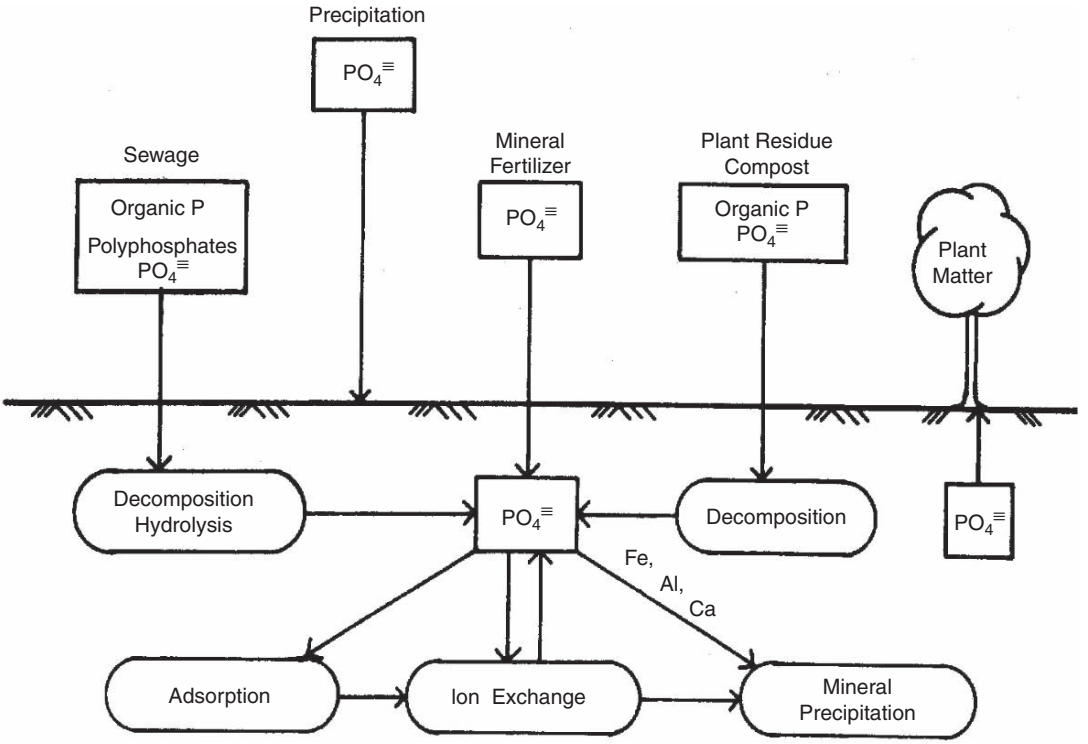

Fig. 2.5. Phosphorus transformations in soil.

3.5.4. Organic Matter Removal

Organic matter in sewage may occur as both living cell material and dead material. The living material is usually bacteria, but may include fungi, algae, and so on. The dead organic matter may consist of food or garbage, waste products from living plants and animals, and dead plants and animals. In general, the organic matter consists of proteinaceous material. Other specific organic compounds that may gain access to the soil are insecticides, pesticides, and petroleum products (55).

Large organic molecules are broken down by hydrolytic enzymes in the soil under both aerobic and anaerobic conditions. Under aerobic conditions within the soil, these smaller organic molecules are used as a source of energy for bacteria that convert some of the organic matter to carbon dioxide and water and some to cell matter. Under anaerobic conditions, the bacteria reduce the organic matter to less complex organic compounds, methane, CO_2, and some cell matter. The aerobic transformation is a more efficient use of energy and results in a greater breakdown of the organic matter to inorganic material (47). It is usually considered more desirable to maintain aerobic conditions within the soil. The breakdown of organic matter requires the presence of oxygen in some form. Normally, this oxygen is obtained from the dissolved oxygen in the applied wastewater and from the air and the soil voids. However,

when too much organic matter is present, the oxygen demand is greater than the oxygen available. Thus the next source of oxygen (an electron acceptor) is that from nitrates and sulfates in the water. Whereas the reduction of nitrates results in the liberation of nitrogen gas, the reduction of sulfates results in the production of highly toxic hydrogen sulfide gas. Thus it is desirable to limit the amount of organic matter added to the soil to prevent the formation of anaerobic conditions. This may be accomplished by applying less total organic matter to the soil, by intermittent dosage to allow oxygen to enter the soil void spaces, and by lessening the concentration of organic material in the applied liquid. Given sufficient time, all of the degradable organic matter will be converted to inorganic materials.

3.5.5. Trace Metals Removal

Trace metals are quite ubiquitous. They are present in the food we eat, the liquids we drink, the wastes we produce, and the fuels we burn. Thus, they may be present in domestic wastes as well as in rainfall. However, they become a significant problem as a result of certain industrial wastes. Concentrations of little more than trace amounts can result in the poisoning of biological systems (39, 55). Because biological systems play such an important part in recycling substances, it is desirable to keep the concentrations of trace metals as low as possible.

Trace metals are adsorbed in the soil and taken up by biological systems (36, 64). Very minute quantities of copper, zinc, and boron have been found to be essential for plant growth (65). These and other trace metals may be taken up by plants (43) and concentrated in a certain portion of the plant. If they are concentrated in the portion consumed by animals, an additional concentration factor may occur in the animal. If, in turn, humans consume this animal portion, there may be a resultant significant concentration of heavy metals in humans. Because of the various food chains and concentration factors involved, it is difficult to set a quantitative level of trace metals that should not be exceeded. In general, where trace metals may be present in significant amounts, it is considered judicious to use such wastes for the production of trees and other nonedible products. This will result in less possibility of direct recycle to humans. Another alternative is to remove the trace metals before application to the land (62).

3.5.6. Disinfection

An unresolved concern is the necessity for chlorination of the waste before irrigation. Whenever there is the possibility of human contact with the wastewater, chlorination is highly recommended to control pathogenic organisms (32). However, much of the purification in soil is carried out by biological agents that might be adversely affected by excessive chlorination. The greatest microbiological activity is near the surface of the soil, which is the portion that would be subject to the highest concentration of chlorine. Furthermore, chlorination adds dissolved solids to the wastewater and ultimately to the soil, and increases the cost of treatment. Some states require chlorination regardless of its necessity. In Michigan (33) chlorination is practiced for spray irrigation, but not for flood irrigation or seepage basins unless underdrains recollect the effluent.

An alternative to chlorination is disinfection using ultra-violet (UV) light. Whereas chlorination requires a residual, UV disinfection produces no residual, thus there is no interference with the desirable organisms responsible for the purification. Similarly ozonation achieves adequate disinfection without producing a residual that would interfere with biological activity in the soil.

Thus, the specific problems to be solved and the conditions under which the irrigation is conducted will govern which system is most applicable for a particular situation.

4. DESIGN

4.1. Preliminary Studies

4.1.1. Infiltration Rates

The rate at which applied water can infiltrate into the ground may be the limiting factor in the design of an infiltrative land application system. There should never be any surface runoff of irrigation water, particularly not to a nearby stream or lake. This prohibition applies at all times of the year, including when the ground is frozen and during spring thaw or a heavy rainfall. For overland flow systems, if the applied irrigation water is not completely infiltrated or evaporated, the residual liquid must be collected for recycle or further treatment. Thus there must always be knowledge of how rapidly applied irrigation water can infiltrate the soil.

This is usually determined by a standard percolation rate test. A hole is dug in the ground with a diameter of approximately 0.3 m (1 ft). The depth to which the hole is dug depends upon the type of irrigation system planned. It must be deep enough to hold the water applied for the test. On the other hand, if surface irrigation is the controlling factor, it may be expedient to build a small wall or dyke around the area where the test is being done to hold the test water. The hole or confined area is then saturated with water. Finally, additional water is added, the level of water is measured, and the liquid is allowed to infiltrate into the soil. After a given period of time, the level is again measured. In performing "percolation" studies for septic tank systems, the common procedure is to determine the time it takes for the water level to fall 1 in. (2.5 cm). Rates are then recorded as min/in. (min/cm). Acceptable infiltration rates may be of the order of 0.5 to 3 cm/h (1/4 to more than 1 in./h). The type of application system must be considered in evaluating the infiltration rate. The infiltration rate is the controlling factor in the ultimate determination of the land area requirements for any land application system.

4.1.2. Land Area Requirements

The amount of land required is a function of the infiltration rate and the amount of water to be applied to the soil (23). The latter, in turn, is determined by the type of application system to be used. Where crop irrigation is practiced, recommended loading depths are in the range of 2 to 10 cm/wk (1 to 4 in./wk) ; this must include natural precipitation. For spraying onto forested land, the application rate may be in the order of 0.5 to 1 m/wk (1.5 to 3 ft/wk). Application rates for rapid infiltration beds may be in the order of 1 or more m/wk (3 ft or more). Frequently, the application rate for infiltration beds is directly controlled by the infiltration rate. The volume of wastewater flow is usually expressed in the depth applied to a

given area per unit of time, usually a week, e.g., ha-cm/wk (ac-in./wk). Using the loading depth in similar units (cm/wk or in/wk) will give the resultant area required in hectares (acres).

4.2. Application Rates

In most situations, intermittent application of the wastewater is desirable. Actual application for irrigation of plants is quite low. In irrigation using rotary sprays, there is an inherent intermittency as the sprayer rotates. The length of time between doses is a function of the soil and the crop cover. Crops should not be allowed to dry to the wilting point. Clay soils should not be allowed to dry completely because they tend to cake, thus preventing further infiltration upon the application of the next dose. Frequency of dosing must be determined on an individual basis considering the type of soil, crop cover, temperature, and precipitation.

In cases where there are no retaining walls, such as surface spreading and spraying, the instantaneous application rates should not exceed the infiltration rate of the soil to prevent surface runoff from the irrigated area. This is no problem where embankments are used to contain the applied water.

For surface application on plant-covered soil, suggested rates of application are about 5 cm (2 in.)/wk (65). However, this is not offered as a firm application rate applicable to all situations. It is suggested as a good starting point, and actual field studies must be made to determine the optimum rate.

For infiltration beds, it is generally recommended that dormant periods be allowed to reaerate the soil (42). The dosing cycle may be long, lasting for periods of from several days to a week or longer (57). Intermittent application is recommended to allow the soil to dry partially so that air may enter into the void spaces to enable aerobic conditions to predominate (31).

The extreme dosing time in intermittent application extends to seasonal usage. Such a regimen is generally coordinated with holding ponds or lagoons. At Sunapee State Park, NH, holding lagoons are provided and spraying takes place annually over about a 6-week period during June and July (59).

4.3. Distribution Facilities

Means must be provided to convey the irrigation water to the site for the land application. Depending upon the elevation of the site, pumping may be required to aid in this conveyance. Duplicate pumping facilities should be provided and in cases where the removal of the wastewater is critical, emergency power supplies must also be provided.

Actual distribution of liquid onto the land may be made through open channels or through various types of pipes. Open channels are acceptable where gravity flow is available. In large irrigation systems, there may be a main channel with several branch channels. Siphons may be used to divert the wastewater from the distribution channels to the fields to be irrigated. Open channels may also serve infiltration beds. In all instances, adequate piping and valving must be provided to divert the flow from one area to another. Where infrequent dosing is to be

used, a holding tank or storage lagoon will be required. This will be drained during the period of application.

For a spray system, a pressure pipe able to maintain the maximum pressure for efficient nozzle operation is required. The pumps must be of capacity to provide the required head at the nozzle plus overcoming the friction within the pipe (24).

The materials used for the distribution system may also be varied (24). Open channels may be of earth with a plastic liner or concrete. Half-sections of inexpensive or lightweight pipe may also be used. For spray systems and where gravity flow cannot be used, piping must be provided to convey the wastewater to the point of application, remembering that the pipe must be of sufficient strength to withstand the pumping pressure required at the nozzle plus the headloss in the pipe. The pipe may be buried or placed on the surface of the ground. In cold climates, provisions must be made for complete drainage during periods of freezing. It is also possible to use lightweight portable and/or flexible piping. In small installations, it may be more economical to buy a limited amount of pipe and move it on a daily basis to the individual sites to be irrigated. In general, at larger installations, more permanent pipe is desired. However, this must be equipped with the proper valving so that the discharged water may be diverted to the location to be irrigated.

4.4. Monitoring

Provisions must be made for proper monitoring to guard against undue contamination of groundwater supplies in the area near the land application system. This will provide a check on the system and will indicate whether or not the system may have to be modified or even terminated because of potential contamination of water supplies in the area (66). Monitoring may be a significant portion of the cost of a land application system (67).

4.4.1. Location

There should be a minimum of two observation wells for any system (65). One observation well should be upstream and the other downstream from the site of the application. Resistivity surveys may be used to locate the waste stream in the ground (28, 68). This will provide information on the natural quality of the ground water to identify any potential pollution or contamination caused by the land application. Obviously, these wells must reach down into the ground water and may be relatively deep depending upon the depth to ground water at the location. In large installations, more than one downstream well may have to be installed. A recommended location of additional observation wells would be between the application site and any nearby water supply.

4.4.2. Tests

A variety of tests should be run on the upstream and downstream monitoring samples. The actual tests performed will be a function of what is desired to be evaluated. This includes the improvement of quality of the applied wastewater, the effect upon the soil itself, and potential contamination of the ground water.

Chlorides should be determined to assure that there is no buildup of this ion in the soil. A buildup could inhibit future crop growth (23).

Organic nitrogen, ammonia nitrogen, and nitrate nitrogen should be determined to assure that all of the reduced nitrogen is oxidized to nitrate and to be certain that this nitrate concentration does not exceed drinking water standards in any water supply. Excess nitrates could also stimulate algal growth in any surface water that the ground water may ultimately reach.

A test for BOD or COD will assure that there is sufficient removal of organic materials by the soil system.

Coliform and fecal coliform analyses should be performed to assure that there is no contamination due to pathogenic organisms.

Orthophosphate and total phosphates should be determined to assure that these are being taken up by the soil and plants and not carried to some surface water-course where they may stimulate algae and aquatic plant growth.

There is less need today to determine detergent concentrations in the soil because of the conversion to biodegradable detergents. However, it is recommended that some check on this substance should be made.

Pesticides, herbicides, and heavy metals may create a potential problem. If it is determined that these are normally absent from the wastewater, it may not be necessary to measure their concentrations in the soil. On the other hand, if there is any potential for these substances to be present, they should be monitored in the downstream ground water. Likewise, if an industrial waste containing a relatively high concentration of any hazardous substance should be present, that substance should be monitored in the ground water.

The frequency of testing is generally a function of the size of the system, the rate of flow through the soil, and the potential for polluting a local ground water. In general, the larger the installation, the more frequent should be the sampling. A generally recommended frequency of sampling is in the range of weekly to monthly during operation of the system.

5. EVALUATION

5.1. Effectiveness

Many years of use of land application systems has shown that this is an effective means of disposing of wastewater and purifying it at the same time (27, 52, 53). Various systems are available, each of which has its own optimum application. For removal of nutrients, particularly nitrogen, application to the root system of a cover crop is most desirable. Where larger volumes of water are to be disposed of and a forested area is available, spraying onto forest areas has been found to be highly successful. Beneficial effects upon wildlife in irrigated forest areas have also been shown (21, 69).

If the prime concern is to dispose of large quantities of water or to replenish the ground-water supply, the use of seepage beds with maximum rates of infiltration has been found to be successful. Phosphorus appears to be removed by such systems, but the nitrogenous compounds converted to nitrate will pass through the soil (52). Nitrate may be taken up by crops and trees, but there is little uptake of nutrients by crops in temperature climates where the crops lie dormant during the winter period. However, evidence has shown that where there is sufficient plant litter on the soil, there is some uptake and storage of the nutrients during the

winter until the time of plant growth. Forested areas appear to have advantages in this respect in that the roots and forest litter tend to keep the soil open even in the winter and there is marked storage of the nutrients during the dormant period for use during the growing season. Furthermore, if heavy metals are present, these may be taken up by the trees and stored in the wood rather than recycled to a food supply system. Type of soil and distance to the nearest water supply are important factors to be considered from the standpoint of potential pollution of groundwater supplies. With judicious awareness of the potential problems, the use of land application for disposal and renovation of wastewaters has been shown to be quite effective.

5.2. Applicability

5.2.1. Scale

One of the concerns about the use of land application for disposal of wastewater is the amount of land required; obviously, this depends upon the type of system employed. The lowest application rate is in the order of 5 cm/wk (2 in./wk) recommended for crop irrigation systems. This results in approximately 0.5 hectares (1.3 acres) for every 100 persons contributing wastewater to the system. Use of forest spraying and rapid infiltration-percolation techniques will require less total area for disposal and, therefore, be more practical for larger population areas, assuming the appropriate type of soil is available.

5.2.2. Location

5.2.2.1. RURAL

The disposal of wastewater by land application is most attractive in rural and suburban areas. A significant amount of land will be required for the disposal site. Furthermore, the right type of soil must be available. Mountainous areas having steep slopes causing rapid surface runoff are generally not recommended for any of these systems. Soil with a reasonable infiltration rate is required. Where spraying onto forested lands is to be considered, the forested lands must be available. Rapid infiltration beds require relatively large areas of level ground, although terracing may be employed. Consideration must be made of the reduction in nutrient removal during the winter period where crop irrigation is employed. In summer recreation areas, this may be sufficient. However, experience has shown that with the increased interest in skiing and other winter sports, many summer recreation homes are being converted to winter hideouts as well. Many factors enter into the consideration of the location of the disposal site and the situation under which the system is applicable.

5.2.2.2. URBAN

Urban areas present a greater challenge for the disposal of wastewater by land application. Some wastewaters may be used for irrigation of park areas within the city or for golf courses in the surrounding areas. However, where land is not available nearby, long pipelines will have to be constructed to convey the liquid to a suitable disposal site. This of course is subject to the usual land value constraints; land is often much more valuable as a housing location than as a waste disposal site. Thus land application is less suitable to large urban areas.

5.3. Cost

Evaluating the true cost of waste treatment by land application is difficult. Each different method of land disposal has a somewhat different cost. Thus an overall estimate would not be truly representative. One of the major expenses for any of these systems is the cost of the land involved (31). Land costs vary from area to area and must be determined on an individual basis. Prime land need not be used for the disposal system; however, the value of land may change with time (31). In general, per capita costs decrease as the population served increases (33). Thus the actual cost of a system must be determined for each situation.

Some of the factors to be considered in evaluating the cost are listed here (70, 71). Knowing the number of persons being served by a system, an estimate can be made of the amount of land required. From information on cost of land in the area, the cost of the land can be calculated. Another significant cost is the distribution system. Where gravity can be employed, this becomes relatively inexpensive. In areas where freezing does not occur, there is less need for caution in setting the grade of pipes, because drainage before freezing temperatures will not be necessary. The type of pipe employed will be a function of whether permanent or portable piping is used and the total pressure against which pumping is required, including both friction loss and the pressure required at a spray nozzle. Pumping can add a significant cost to the disposal system (72). Valves must be installed at appropriate locations to assure the proper distribution of the water for disposal. Distance from the source of the waste to the application site obviously is an important factor. Allowance must be made for the cost of a monitoring system and the analyses required to assure the maintenance of high water quality standards.

An important consideration is the degree of pretreatment before the actual land application process. In general, land application systems are designed for final treatment of the water. In most instances today, this precludes the equivalent of biological secondary treatment of the wastewater. There are a few locations where only primary settled effluent has been discharged onto the ground, producing satisfactory treatment (1, 18). The Bolton Landing, NY, infiltration-percolation treatment plant bypasses the trickling filters during the winter because of icing problems (73), and the Ft. Devens, MA, rapid infiltration plant has been successfully treating primary settled effluent since 1942 (74, 75). Also, it is US EPA's policy under PL95-217 not to fund secondary treatment before land application unless it has been proven to be necessary. On the other hand, septic tank effluents must be considered a form of land application. Septic tanks are very inexpensive to operate. Thus the cost for disposal by irrigation should be compared with that of tertiary treatment of wastewaters. In general, the cost of tertiary treatment by land application techniques is far less than by any other commonly used tertiary treatment system (28, 29).

Evaluation of the net cost of a land application system for wastewater disposal must include potential sources of income as a result of or in conjunction with the system. These may include the following (71): (a) increased agricultural production, (b) increased wildlife production (21, 69), (c) long-term increase in land values, (d) isolation areas for future (nuclear) electric power stations, (e) sale of cooling water, (f) sale of industrial process water, (g) associated solid waste processing facility, (h) use of the area for recreational purposes, particularly during

the nonirrigation season, as in winter (59), (i) reclamation of land such as strip-mined areas (19, 76), and (j) reduction of fire hazards in drought afflicted areas (77).

A study (78) showed that, on the average, disposal on land is less costly than the equivalent conventional techniques. There is a greater saving in cost at a small treatment plant than at a larger treatment plant. Furthermore, the greater the required BOD reduction, the greater is the cost advantage of the land treatment system. A direct comparison is difficult, because means are not available for achieving a predetermined degree of treatment in a land disposal system. The degree of treatment for all land application techniques approaches 100% BOD removal. Thus in general, irrigation techniques achieve a greater degree of treatment. The report indicated that for 85% BOD removal, land disposal for a 0.5 mgd (2000 m^3/d) plant can effect a saving of 7¢/1000 gal (2¢/m^3) treated, whereas a 10 mgd (38,000 m^3/d) plant will save approximately 5¢/1000 gal (1.3¢/m^3). For 95% BOD removal, average savings for a 0.5 mgd (2000 m^3/d) plant would be in the order of 40¢/1000 gal (10¢/m^3), and for a 10 mgd (38, 000 m^3/d) plant, approximately 14¢/1000 gal (4¢/m^3). Additionally, savings of 7 to 15¢/1000 gal (2 to 4¢/m^3) could be anticipated from the sale of byproducts (irrigation water and nutrients). The report concluded that, in general, municipalities could anticipate a savings in treatment costs by using land disposal techniques where possible.

In a study comparing the actual energy requirements for operation of the Lake George Village rapid infiltration plant with the proposed (equivalent size) activated sludge plant, it was estimated that the activated sludge system would consume 10 times as much (electrical) energy as the existing land application system including its trickling filter system (79).

A detailed analysis of the costs for land, preapplication treatment, transmission, storage, and land application, and recovery of renovated water has been prepared for the US EPA (80). For preliminary screening costs, curves are presented for capital costs, amortized costs, and operation and maintenance costs at flows from 4.38 to 4380 L/s (0.1 to 100 mgd). Cost calculation procedures and an illustrative example are included. For detailed planning costs, useful curves, tables, and data are presented for 33 individual components related to either flow rate or field area. For capital items, total construction costs are shown, and operation and maintenance costs are divided into labor, materials, and power where applicable.

One of the most difficult costs to evaluate will be the monitoring costs. It appears that US EPA regulations for best practical treatment will require that for infiltration systems the groundwater quality must be maintained at drinking water standards. This will require monitoring for trace organics, heavy metals, and other trace substances in the ground water. The costs for this monitoring could be such that disposal by another method not requiring such stringent monitoring might become less expensive, thus deterring the use of this otherwise desirable disposal system.

5.4. Ease of Design for Various Conditions

The final design and selection of equipment for a land application system must be conducted by a competent engineer. However, certain guidelines are offered to assist anyone in evaluating a system and estimating the amount of land required for a given system. The most important parameters to be measured are: (a) the amount of water to be disposed of, (b) the weekly loading rate to be applied, which is a function of the type of system to be used, and

(c) the hourly application rate, which is a function of the infiltration rate of the soil to be used. It immediately becomes obvious that each system must be evaluated on its own basis.

5.4.1. Water Flow

The daily water flow to be disposed of should be measured where possible or estimated on the basis of the type of population. Where no other information is available, an average per capita flow of $0.4\,m^3/d$ ($\sim$100 gpd) may be used. For convenience in estimating the area required, this should be converted to hectare-centimeters per week (acre-inches/week). This may be done as follows:

$$\text{ha} - \text{cm/wk} = \frac{\text{m}^3/\text{d} \times 7\text{d/wk}}{100\,\text{m}^3/\text{ha} - \text{cm}} \text{ or } \frac{\text{L/d} \times 7\text{d/wk}}{10^5\,\text{L/ha} - \text{cm}} \tag{6}$$

or

$$\text{ac} - \text{in/wk} = \frac{\text{gal/d} \times 7\text{d/wk}}{27{,}154\,\text{gal/ac} - \text{in}} \tag{7}$$

5.4.2. Loading Depth

Loading depth represents the depth of liquid applied to the area in use over a unit of time, conveniently, a week. This is used because a low application rate of about 5 cm/wk (2 in./wk) is suggested to allow complete uptake of the applied water by plants (24). This value represents the lowest loading depth normally encountered. The other extreme is in terms of m/wk (ft/wk) for recharge of ground water or disposal of large amounts of water by rapid infiltration technique. In this instance, the soil infiltration rate may possibly be the maximum loading depth allowable. Continuous application may be conducted; however, rest periods are recommended.

5.4.3. Application Rate

The instantaneous application rate is a function of the infiltration rate of the water into the soil (24). It is usually expressed as cm/h (in./h) (although it is sometimes measured as min/in (min/cm)) and is calculated as described in Section 4.1.1. Application in excess of the infiltration rate should not be practiced in open fields, otherwise surface runoff will occur. Where embankments are used to contain the liquid, the temporary amount of water placed in the beds may exceed this rate, but over an extended period of time, the total application rate cannot exceed the infiltration rate. This value must be measured in the field for every situation.

5.4.4. Area Required

Knowing the amount of water to be disposed of and the loading permitted in terms of either loading depth or application rate, the land area required can be calculated as follows:

$$\text{Area, ha} = \frac{\text{flow, ha} - \text{cm/wk}}{\text{depth, cm/wk}} \tag{8}$$

or

$$\text{Area, ac} = \frac{\text{flow, ac} - \text{in/wk}}{\text{depth, in/wk}} \tag{9}$$

In general, the limit for crop irrigation is the amount of water applied so that it is all taken up by the crops. For groundwater recharge, the limit approaches the infiltration rate with allowance for resting time. Thus each system must be considered individually.

Additional factors that must be considered in the hydraulic loading include precipitation, evapotranspiration, percolation, and possibly runoff. An important factor in designing for a crop irrigation system is that the crops may have a short growing season depending upon location. Storage may have to be provided during the winter when irrigation is not practiced.

Factors other than hydraulic loading may limit the loading rate. These include nitrogen, phosphorus, organic matter, cation exchange capacity (CEC), or other constituents. On the basis of balances of these substances, a loading rate can be established for each parameter, and the land area required calculated for each. The factor requiring the largest land area dictates the final design size of the system. Examples of calculations for the land requirements for some of these parameters are presented for both the irrigation and the rapid infiltration systems.

5.4.5. Sample Calculations

5.4.5.1. Per Capita Loadings

A typical calculation for design of a land application system is offered on the basis of a population contribution of 1000 persons. This is a convenient number that can be translated into other size populations with relative ease. Per capita wastewater flows are assumed to be approximately $400\,\mathrm{L/d} = 0.4\,\mathrm{m^3/d}$ (100 gpd). Calculation of the weekly flow is shown below:

$$\frac{1000\,\text{persons} \times 0.4\,\mathrm{m^3/person - d} \times 7\mathrm{d/wk}}{100\,\mathrm{m^3/ha - cm}} = 28\,\mathrm{ha - cm/wk} \tag{10}$$

or

$$\frac{1000\,\text{persons} \times 400\,\mathrm{L/person - d} \times 7\mathrm{d/wk}}{10^5\,\mathrm{L/ha - cm}} = 28\,\mathrm{ha - cm/wk} \tag{11}$$

or

$$\frac{1000\,\text{persons} \times 100\,\mathrm{gal/cap - d} \times 7\mathrm{d/wk}}{27{,}154\,\mathrm{gal/ac - in}} = 25.77\,\mathrm{ac - in/wk} \tag{12}$$

Because the loading depth will vary depending on the type of system used, calculations are offered for the two typical types of system used.

5.4.5.1.1. Crop irrigation For crop irrigation, the limit of loading depth is normally the application rate that permits all of the liquid to be taken up by the roots. This is in the order of, but not exclusively, 5 cm/wk (2 in./wk) (65). Calculation for the area required for 1000 persons would be as follows:

$$\frac{28\,\mathrm{ha\text{-}cm/wk}}{5\,\mathrm{cm/wk}} = 5.6\,\mathrm{ha} \qquad \frac{25.77\,\mathrm{ac\text{-}in/wk}}{2\,\mathrm{in/wk}} = 12.9\,\mathrm{ac}$$

5.4.5.1.2. Rapid infiltration-percolation For groundwater recharge, the limiting rate of application is the infiltration rate of the soil. Although continuous application may be used, in most instances, the application is intermittent on a weekly or monthly basis. Therefore, some

additional area needs to be provided for the discontinuance of application. Usually, an open sandy or gravelly soil is used for such application and the infiltration rates are relatively rapid. In these instances, the loading depth is frequently in the order of several meters per week depending upon the soil being dosed. A typical value may be in the range of 1 cm/h (about 0.4 in./h) (65). This value is used merely for calculation; the actual value would have to be determined for each individual area. At a rate of 1 cm/h, the weekly application rate would be calculated as follows:

$$1\,\text{cm/h} \times 24\,\text{h/d} \times 7\,\text{d/wk} = 168\,\text{cm/wk}$$
$$0.4\,\text{in/h} \times 24\,\text{h/d} \times 7\,\text{d/wk} = 67.2\,\text{in/wk}$$

The land area required per 1000 persons would then be calculated based on the wastewater flow and the application rate as follows:

$$\frac{28\,\text{ha-cm/wk}}{168\,\text{cm/wk}} = 0.17\,\text{ha} \qquad \frac{25.77\,\text{ac-in/wk}}{67.2\,\text{in/wk}} = 0.38\,\text{ac}$$

As a rule of thumb, this area is frequently doubled to allow for resting periods.

5.4.5.2. LIQUID LOADING BASED ON WATER BALANCE (80)

The factors considered in the water balance include the amount of effluent applied, the precipitation, evapotranspiration, percolation, and runoff.

5.4.5.2.1. Crop irrigation For irrigation systems, the water balance is based upon the following equation:

$$\text{Effluent applied} + \text{precipitation} = \text{evapotranspiration} + \text{percolation} + \text{runoff} \qquad (13)$$

Percolation should be minimal or none at all. Runoff may be considered to maintain a salt balance in the soil, but is generally none. If there is runoff, it must be collected and controlled.

Seasonal variations in each of the above values should be taken into account. This may be done by means of a spreadsheet to evaluate the water balance for each month as well as the annual balance.

The design value for precipitation should be determined on the basis of a frequency analysis of wetter than normal years. The wettest year in 10 is suggested as reasonable in most cases; however, it is prudent to check the water balance using the range of precipitation amounts that may be encountered. For purposes of evaluating monthly water balances, the design annual precipitation can often be distributed over the year by means of the average distribution, which is the average percentage of the total annual precipitation that occurs in each month. Again, the range of monthly values that may be encountered should be analyzed, especially for the months when the storage reservoir is full.

Evapotranspiration will also vary from month to month; however, the total for the year should be relatively constant.

When irrigating in arid climates, it is necessary to remove the salts that accumulate in the root zone as a result of evaporation. Some amount of percolation or runoff may be necessary to accomplish this leaching.

Table 2.1
Balance sheet of evapotranspiration

	Water losses			Water applied		
Month (1)	Evapotranspiration, cm (2)	Percolation, cm (3)	Total, cm (4)	Precipitation, cm (5)	Effluent applied, cm (4) − (5) = (6)	Total, cm (5) + (6) = (7)
Jan.	1.9	10.0	11.9	7.1	4.8	11.9
Feb.	3.8	10.0	13.8	7.2	6.6	13.8
Mar.	7.8	10.0	17.8	8.9	8.9	17.8
Apr.	10.0	10.0	20.0	10.6	9.4	20.0
May	13.3	10.0	23.3	8.1	15.2	23.3
Jun.	16.5	10.0	26.5	6.6	19.9	26.5
Jul.	17.8	10.0	27.8	3.5	24.3	27.8
Aug.	16.5	10.0	26.5	2.3	24.2	26.5
Sep.	11.4	10.0	21.4	4.1	17.3	21.4
Oct.	10.0	10.0	20.0	8.6	11.4	20.0
Nov.	3.8	10.0	13.8	5.8	8.0	13.8
Dec.	2.0	10.0	12.0	7.8	4.2	12.0
Total annual	114.8	120.0	234.8	80.6	154.2	234.8

The amount of effluent that may be applied without causing runoff, but still allowing some predetermined degree of percolation (10 cm/mo, for this example) may be determined using a spread sheet as indicated in Table 2.1. It is assumed that the soil infiltration rate is greater than 10 cm/mo., and that the instantaneous application rate is less than the infiltration rate, otherwise runoff will occur. Evapotranspiration is a function of temperature, latitude, and crop cover.

Based on the calculations shown I Table 2.1, the total liquid that can be applied is 154.2 cm. From this, the average weekly flow is $154.2/52 = 3.0\,\text{cm/wk}$, and the average monthly flow is 12.8 cm/mo. If the application must be kept constant, the increased total flow during rainy months would increase the total percolation or result in runoff if the total flow exceeds the infiltration rate. To prevent increased percolation or runoff, the application rate would have to be decreased. Storage would have to be provided when the flow exceeds the determined loading. By calculating the cumulative excess of flow, the total maximum storage requirements may be determined. The alternative is to increase the land area used.

5.4.5.2.2. Rapid infiltration-percolation For groundwater recharge, the water balance is based upon the following equation:

$$\text{Effluent applied} + \text{precipitation} = \text{evaporation} + \text{percolation} \tag{14}$$

There is no runoff in this system. The percolation rate used should be that under saturated conditions, because this condition frequently occurs and is the condition during which the

infiltration rate is minimal and becomes limiting. Also allowance must be made for clogging of the soil surface after extended use.

For higher rate systems and systems with intermittent applications, percolation is the major factor, with evaporation accounting for 10% or less of the effluent applied. Precipitation may be significant in humid climates and is accounted for in the same manner as for irrigation, using a frequency analysis of the available data. In arid climates, the precipitation should not be omitted, because it often occurs in large amounts during a few months.

Calculation of the amount of effluent that may be applied in a recharge system may be determined using a spread sheet similar to that used to determine the applied rate for crop irrigation (Section 5.4.5.2.1.).

5.4.5.3. Loading Based on Nitrogen Balance (80)

A total nitrogen balance can be as important as a water balance, because nitrate ions are mobile in the soil and can affect the quality of the ground water. On an annual basis, the applied nitrogen must be accounted for in crop uptake, denitrification, volatilization, addition to ground water or surface water, or storage in the soil.

Calculation of the allowable loading must be based upon total nitrogen, because organic, ammonia, nitrite, and nitrate all interact in the soil. The total nitrogen loading to the soil may be calculated by the equation:

$$N = 0.1CL \tag{15}$$

where N = annual nitrogen loading kg/ha-yr, C = total nitrogen concentration, mg/L; and L = annual liquid loading, cm/yr. Or

$$N = 2.7CL \tag{16}$$

where N = annual nitrogen loading, lb/ac-yr; C = total nitrogen concentration, mg/L; and L = annual liquid loading, ft/yr.

The crop uptake of nitrogen is dependent upon the crop grown, the pH of the soil, the nitrogen starvation of the crop, and the amount of nitrogen available in the irrigation water. Ranges of nitrogen uptake from fresh water are from 55 to 675 kg N/ha-yr (50 to 600 lb N/ac-yr). Unpublished data (81) using primary and secondary sewage treatment plant effluent have shown the nitrogen uptake by forage to be in the range of 389 to 542 kg N/ha-yr. When more than one crop per year is grown on the same field, the total nitrogen uptake for the entire year should be determined. Nitrogen removal by crop uptake is a function of crop yield and requires the harvesting and physical removal of the crop to be effective.

The extent of denitrification and volatilization depends on the loading rate and characteristics of the wastewater to be applied, and the microbiological conditions in the active zones of the soil. Volatilization of ammonia will not be significant for effluents with a pH less than 7 or for nitrified effluents. For irrigation systems, denitrification is generally of minor importance, depending upon the soil, the application rate, and the crop. Denitrification may be a significant nitrogen removal mechanism for overland flow systems because observed removals cannot be accounted for solely by crop uptake. For high-rate

infiltration-percolation systems, denitrification is the only significant mechanism of nitrogen removal from the system. By managing the hydraulic loading cycle to create alternately anaerobic and aerobic conditions, Bouwer (82) obtained up to 80% nitrogen removal as a combined result of ammonia adsorption and denitrification during most of the period of inundation. Over a 4-yr period the calculated removal was 30% at a loading rate of 21,000 lb/ac-yr (23,450 kg/ha-yr). Without special management techniques, overall nitrogen removal may be only 10% or less.

The soil mantle cannot hold nitrogen indefinitely, although organic nitrogen can be stored in the soil to a certain extent. The ammonium and organic nitrogen is ultimately converted to nitrate nitrogen, which can leach out of the soil. Unless nitrogen is taken up by crops and physically removed by harvesting, or the nitrates are converted to nitrogen gas by denitrification, the nitrogen will appear eventually in the runoff or percolate.

It may be seen that establishing a nitrogen balance is more difficult and less reliable than calculating the liquid loading. Much more data must be obtained, and that which is available is less precise. There are many variables that are not even considered. Thus a nitrogen balance is at best a rough estimate.

5.4.5.4. Loading Based on Phosphorus (80)

Phosphorus is removed from percolating wastewater by fixation and chemical precipitation. For irrigation, the phosphorus loading will usually be well below the capacity of the soil to fix and precipitate the phosphorus. Typically, the crop uses less than 20% of the phosphorus applied and the remainder stays in the topsoil. Soil column tests are frequently conducted to determine the fixation capacities of the soil; however, the results of these tests should be used with caution because long-term behavior and the effects of time cannot be duplicated in a short-term test.

For infiltration-percolation systems, fixation and chemical precipitation in the soil are responsible for phosphorus removal. As with irrigation, the capacity of the soil to remove phosphorus can be estimated from laboratory tests. This capacity can be quite high even for sandy soils with relatively low fixation capacities. Greater than 99% phosphorus removal has been reported in sand at Lake George with flow through approximately 600 m (2000 ft) of sand (3). These results also show that laboratory absorption tests cannot be relied upon completely to estimate total phosphate uptake capacity of a soil, because such studies (28) predicted phosphorus breakthrough in about 10 yr, whereas the system at Lake George is still performing satisfactorily since 1939.

5.4.5.5. Loading Based on Organic Matter (80)

The average daily organic loading rate may be calculated from the liquid loading rate and the BOD concentration of the applied effluent. Between 10 and 25 lb/ac-d (11 and 28 kg/ha-d) are needed to maintain a static organic-matter content in the soil. Additions of organic matter at these rates help to maintain the tilth of the soil and replenish the carbon oxidized by microorganisms, and would not be expected to pose problems of soil clogging. Higher loading rates can be managed, depending upon the type of system and the resting period. Based upon 10 to 25 lb/ac-d (11 to 28 kg/ha-d) of BOD, the resultant addition of 2 lb/ac-d (2.2 kg/ha-d)

or less from a typical secondary effluent applied for irrigation should not pose a problem of organic buildup in the soil. When primary effluent is used, organic loading rates may exceed 20 lb/ac-d (22 kg/ha-d) without causing problems.

Resting periods are standard with most irrigation techniques to give soil bacteria time to break down organic matter and allow the water to drain from the soil. Aerobic conditions are thus restored as air penetrates into the soil. Resting periods for spray irrigation may range from less than 1 to 14 days, with 5 to 10 days being common. The resting period for surface irrigation can be as long as 6 weeks, but is usually between 6 and 14 days. The resting period depends upon the crop, the number of individual plots in the rotation cycle, and management considerations.

Organic loading is an important criterion for rapid infiltration systems, because it is related to the development of anaerobic conditions. To meet the oxygen demand created by the decomposing organic and nitrogenous material, an intermittent loading schedule is required. This allows air to penetrate the soil and supplies oxygen to the bacteria that oxidize the organic matter and ammonium. Bouwer (82) reports BOD loadings of 45 lb/ac-d (50 kg/ha-d) using secondary effluent and a liquid loading of 300 ft/yr (91 m/yr). The application cycle consisted of loading for 14 days, followed by 10 days of resting in the summer and 20 days of resting in the winter.

5.4.5.6. Loading Based on Heavy Metals

Heavy metals may be of concern when industrial wastes containing heavy metals or sewage sludges are applied to the soil. Of prime concern is the transport of the heavy metals to the edible portion of the plant. Whereas this involves many factors, a simplified means of estimating the amount of sludge that may be applied has been proposed (83) based on the soil's cation exchange capacity (CEC) and the concentration of zinc, copper, and nickel in applied sludge as follows:

$$\text{Ton, T (metric) dry solids/ha} = \frac{73{,}024 \times CEC}{\text{mg Zn/L} + 2\,(\text{mg Cu/L}) + 4\,(\text{mg Ni/L}) - 30} \tag{17}$$

or

$$\text{ton, t (Eng.) dry solids/ac} = \frac{32{,}600 \times CEC}{\text{mg Zn/L} + 2\,(\text{mg Cu/L}) + 4\,(\text{mg Ni/L}) - 30} \tag{18}$$

in which CEC = cation exchange capacity of soil, meq/100 g dry soil. Based on the quantity of the zinc, copper, and nickel, the land area required can be determined. Restricting loading to less than this calculated limit is expected to prevent the buildup of these and other normally related heavy metals in the soil.

Loading limits for additional metals based upon application of sewage sludges have been specified in 40 CFR Part 503. Table 2.2 summarizes Tables 2 and 4 from Section 503.13 (84). The annual application limits are based on a 20-year life of a disposal site with no attenuation of these metals during this lifetime period. Knowing the concentrations of these metals in the applied wastewater and the volume of liquid, the loadings can be determined. These

Table 2.2
Limits of metals for land application of sewage sludge (84)

Metal	Cumulative loading kg/ha	Annual loading kg/ha-yr
Arsenic	41	2.0
Cadmium	39	1.9
Chromium	3,000	150
Copper	1,500	75
Lead	300	15
Mercury	17	0.85
Nickel	420	21
Selenium	100	5.0
Zinc	2,800	140

limits are designed for sludge loadings, and are seldom approached in wastewater application systems.

5.4.5.7. Loading Based on Suspended and Dissolved Solids (80)

High concentrations of suspended solids such as are found in raw sewage can clog the components of a distribution system and reduce the infiltration rate into the soil. As a result, preapplication treatment for suspended solids reduction may be necessary. The organic fraction of the suspended solids applied to the land is degraded as described for BOD. The inorganic or mineral fraction of the suspended solids is filtered and becomes incorporated into the soil.

Dissolved solids in wastewater may be classified by the extent of their movement through the soil. Chlorides, sulfates, nitrates, and bicarbonates move relatively easily through most soils with the percolating water. These compounds can therefore be leached with applications of wastewater or with rainfall.

Other dissolved solids, such as sodium, potassium, calcium, and magnesium, are exchangeable and react within the soil so that their concentrations in the percolating water will change with depth. Other constituents, such as heavy metals, boron, fluoride, and other trace elements or pesticides, may or may not be removed by the soil matrix, depending upon such factors as clay content, soil pH, and soil chemical balance. On the basis of the analysis of wastewater characteristics and the requirements for groundwater protection, any constituent suspected of having a limiting loading rate should be identified. The loading rate of that constituent should then be calculated, and the resulting land requirement should be calculated.

5.4.5.8. Land Area Requirements Based on Other Than Hydraulic Loading

From the nonhydraulic loadings, as described in Sections 5.4.5.3–5.4.5.7 above, land area requirements may be determined from the following equation:

$$A = \frac{3.65 \times 10^{-5} \times CQ}{Lp} \tag{19}$$

where A = area, ha; C = concentration of constituent, mg/L; Q = flowrate, L/d; Lp = permissible loading, kg/ha-yr, or

$$A = \frac{3040\ CQ}{Lp} \tag{20}$$

where A = area, ac; C = concentration of constituent, mg/L; Q = flowrate, mgd; Lp = permissible loading, lb/ac-yr.

If the permissible loading, Lp, is less than the actual loading (N, as described in Section 5.4.5.3 above), then the area required based upon the permissible loading becomes the limiting factor. The constituent requiring the greatest area based upon the permissible loading will become the limiting constituent. The largest area calculated would be the one used for the design of the system.

5.4.5.9. Design of Overland Flow Systems

In the design of an irrigation system, the prime concern is providing sufficient moisture to supply the water needs for crops being grown. The amount of water to be applied is normally controlled by the climate, the precipitation, and the type of crop. In a rapid infiltration system, the prime concern is the removal of the liquid portion of the wastewater with potential recharge of the groundwater.

An overland flow system, however, is designed for the purification of the wastewater as opposed to the disposal of the liquid. Therefore the area needed may be calculated on the basis of the degree of treatment desired. This is usually based upon the removal of BOD, but may also be a function of ammonia oxidation, total suspended solids removal, and phosphorus removal. Overland flow systems are seldom designed for phosphorus removal, as the removal does not appear to be a function of the loading. Average phosphorus removal in an overland system is in the order of 50% (85).

To formulate design parameters based upon treatment in an overland flow system, the system is treated as a biological film reactor, similar to a trickling filter. Thus, kinetic reactions observed in a trickling filter may be applied to the overland flow system. In a trickling filter the degree of treatment is directly related to the hydraulic detention time (86). In an overland flow system, the hydraulic detention time is dependent upon application rate, slope of the terrace, length of the field, surface microtopography, soil infiltration rate, evapotranspiration, climate, and vegetation density. The only factor controllable by the designer is the application rate. All other factors are site specific.

Studies conducted both by CRREL (85) and Smith and Schroeder (86) have indicated that the mean hydraulic detention time can be estimated by the use of the equation

$$\overline{T} = \frac{0.078L}{S^{1/3}q} \tag{21}$$

in which $\overline{T}$ is the average detention time in minutes, L is the length of the terrace in meters, S is the mean slope in meters per meter, and q is the average overland flow rate in m^3/hr-m width of the field.

Experimental data from CRREL and University of California, Davis have shown that removal can be expressed as a first-order equation in the form

$$\% \text{ Removal} = 1 - (Ae^{-k\overline{T}})(100) \tag{22}$$

Observed coefficients for A and k were as follows:

Coefficient	A	k
BOD removal	0.52	0.03/min
Ammonia removal	0.81	0.03/min

Total suspended solids removal appears to be a function of surface contact and is little affected by detention time. Over the range of detention times tested, removal at 20 min was 86%, and removal at 60 min was 92%.

In the design of an overland flow system, consideration must also be made of the loss of water in flowing over the field. The volume of runoff is frequently 60% to 90% of that applied (85). Thus removal efficiency must be calculated on a mass basis rather than on a concentration basis. Knowing the fraction of the applied liquid that appears at the bottom of the field, a percent removal of a constituent may be calculated as follows:

$$\% \text{ Removal} = \frac{(I \times \text{initial concentration}) - (\text{runoff fraction} \times \text{final concentration})}{I \times \text{initial concentration}} \tag{23}$$

where I represents the total initial flow.

Based on a desired percent removal, the average detention time, $\overline{T}$, can be calculated. Using this value of the mean detention time, and a known slope of a given field, a balance can be made between the length of the slope and the average overland flow rate. It must be noted that the average overland flow rate includes a function of the width of the field and this is considered to be a site-controlled factor. In this way, an overland flow system can be designed to produce a desired degree of treatment. Conversely, knowing an existing set of parameters, the degree of treatment to be expected may be estimated.

5.4.5.10. Total Land Area Requirements (80)

The above calculations represent the area required for the actual crop irrigation or infiltration beds only. The total land area required for a complete system includes allowances for pretreatment; buffer zones; storage, if necessary; sites for buildings, roads, and ditches; and land for emergencies or future expansion. If any on-site preapplication treatment, such as screening, sedimentation, biological or chemical treatment, or disinfection, is required, an allowance must be made for the land needed for these facilities.

A distinction should be made between field area and wetted area. Field area represents the area of the treatment system. The wetted area refers to the area to which liquid is directly applied, either the area covered by the diameter of the spray or the area inundated by surface application. The significance of this difference varies with the treatment method. For spray irrigation, the wetted area may vary from 75% to 100% of the field area. The percentage will depend upon the shapes of the fields, the sprinkler discharge patterns, and the degree of spray

overlap. The highest ratio of wetted area to field area (0.95 to 0.99) occurs with flooding and with ridge and furrow systems. The wetted area should be nearly equal to the field area for most infiltration-percolation systems. For constructed spreading basins, considerable land may be lost in the side slopes of the basin levees.

Although there is little actual data concerning aerosols, there is considerable concern about the effects of aerosol-borne pathogens (87). Therefore, application by spraying may require buffer zones or other measures to ensure that aerosols are contained on the site. Buffer zones ranging from 50 to 200 ft (15 to 61 m) wide have been reported, although requirements for even larger buffer zones may exist.

Irrigation and overland flow systems will generally require off-season or winter storage. Storage may also be useful to equalize flow rates or to provide emergency backup. The land required for storage lagoons or ponds may be considerable, especially in cold climates. Infiltration-percolation systems incorporating spreading basins can usually operate throughout the year, if the limiting loading rate was established for winter conditions.

Area for potential future expansion of a land-application system should be considered in the planning stage. If it is known that the adjacent land is planned for development and will be unavailable for future use, the system should not be referred to as a long-term solution. Often, it is prudent to obtain excess land in case of emergency use. Such things as excessive rainfall, breakdown of preapplication treatment operations, or natural disasters would constitute emergencies.

It may be seen that the type of land application system planned plays a major role in the amount of land required for this method of disposal of wastewater. Each potential system must be evaluated on its own merits and an appropriate choice made to satisfy all requirements. Additional technical information on waste treatment by application onto land can be found in the literature (88–106).

NOMENCLATURE

BOD = Biochemical Oxygen Demand
CEC = Cation Exchange Capacity
CFR = Code of Federal Regulations
DO = Dissolved Oxygen
US EPA = United States Environmental Protection Agency
L = Liters (when not part of BOD equation)
T = Ton (Metric)
t = ton (English)

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3
Treatment by Subsurface Application

Nicholas L. Clesceri, Donald B. Aulenbach, and James F. Roetzer

CONTENTS

Abstract There are two major components of a subsurface wastewater treatment system: (a) a pretreatment tank for gravity separation and accumulation of the settleable solids from the wastewater, and (b) subsurface distribution and final treatment of the supernatant liquid from this pretreatment tank. This chapter covers the topics of subsurface waste treatment theory, anaerobic septic tank, aerobic tank, subsurface disposal, leaching field, tile field, soil percolation, seepage pit, cesspool, upflow permeameter, etc. Many design examples are presented.

Key Words Wastewater treatment • subsurface application • septic tank • aerobic tank • tile field • leaching field • soil percolation • seepage bed • cesspool • upflow permeameter • design • construction.

1. INTRODUCTION

Public health considerations demand proper treatment of domestic wastes. Numerous diseases such as typhoid fever, dysentery, and diarrhea are transmitted by fecal contamination of food and water.

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The United States Public Health Service lists six requirements for proper treatment of domestic wastes (1):

1. They will not contaminate any drinking water supply.
2. They will not give rise to a public health hazard by being accessible to insects, rodents, or other possible carriers that may come into contact with food or drinking water.
3. They will not give rise to a public health hazard by being accessible to children.
4. They will not violate laws or regulations governing water pollution or sewage disposal.
5. They will not pollute or contaminate the waters of any bathing beach, shellfish breeding ground, or stream used for public or domestic water supply purposes, or for recreational purposes.
6. They will not give rise to a nuisance due to odor or unsightly appearance.

These requirements can be met by central sewage collection and treatment systems; however, in many areas of low population density, economic factors preclude collection and central treatment of wastes.

Adequate treatment of wastes becomes a matter of serious concern. It is essential that individual household treatment systems are properly designed to achieve the above requirements. The magnitude of the problem becomes apparent when it is considered that about one-third of the individual dwellings in the United States and Canada depend upon individual waste treatment systems, with this trend continuing in new construction (1–3). In most cases, such treatment involves the septic tank, a unit little changed since its introduction in 1881. The nuisances created by failure of some of these units have led to considerable research into the problems of individual treatment systems.

In many suburban areas individual treatment systems can be considered a temporary solution to waste disposal, because increased population usually results in central sewage treatment becoming economical. In many rural areas, however, individual treatment systems must operate on a permanent basis. The design of any individual system must take into account the fact that all maintenance and repair costs are the responsibility of the individual household. Thus, the treatment method chosen must be not only effective in treating the waste, but also extremely reliable and essentially maintenance-free. Proper design of a system can insure these objectives.

2. THEORY

There are two major components of individual household treatment systems. The first part consists of a tank for gravity separation and accumulation of the settleable solids and the grease from the wastewater, as well as some measure of treatment of the liquid. The second part provides a method for subsurface distribution and final treatment of the clarified liquid from this pretreatment tank. Both of these units can be designed either as an aerobic process (possibly with limited oxygen) or as an anaerobic process. The most common system includes the anaerobic septic tank followed by an aerobic leach field.

2.1. Pretreatment in a Tank

2.1.1. Septic Tank

The primary function of the septic tank is the removal of settleable and floatable solids from the wastewater to prevent their clogging the effluent distribution system. The most important

Table 3.1
Degree of treatment in pretreatment tank, typical values (3)

	Raw waste	Septic tank effluent	Aerobic tank effluent
BOD, mg/L	200	180	40
SS, mg/L	300	150	80
DO, mg/L	1	0	3
Coliform, 10^6/100 mL	12	11	5

design parameter is size, i.e., providing sufficient detention time for solids settling. In the operation of the septic tank, settleable solids settle to the bottom of the tank, where they accumulate and are digested anaerobically, with a resultant reduction in solids volume. In addition, a scum layer composed of grease and other lighter-than-water components builds up at the liquid surface in the tank. The detention time of the tank is based upon the clear water space, or volume between the sludge and scum layers. In practice, two-thirds of the tank volume is reserved for sludge and scum accumulation, with the remaining one-third to provide a detention time of 12 to 24 hours. When they fill their allotted volume, the sludge and scum must be removed by pumping out the tank. This generally is required every two to five years, and is determined by inspection. The detention time has been shown to be important in solids removal (4, 5). Essentially no solids removal occurs in a tank whose detention time has been decreased because of too great a sludge accumulation, whereas an increase in detention time results in increased solids removal. A degree of treatment typical of a septic tank is shown in Table 3.1.

The most common form of a septic tank for an individual household is a single compartment tank as shown in Figure 3.1 (6). Generally, these are precast concrete tanks that are transported and placed at locations as needed. Alternatively, they may be constructed in place. Septic tanks may also be made of steel coated with an asphalt or rustproof coating. Acceptable lightweight materials include fiberglass, fiber reinforced plastic, and polyethylene. As may be seen in the figure, the inlet pipe is baffled so that the incoming sewage can be directed downward, thereby forcing the solids to continue on downward to the bottom of the tank. Similarly, there is an outlet baffle to prevent the scum layer from passing out into the effluent. The solids should be removed from the tank before the scum buildup reaches the bottom of the outlet pipe or before the sludge buildup increases to this same level.

For larger houses or for units of several houses using one septic tank, a larger multicompartment tank is recommended. A typical two-compartment tank is shown in Figure 3.2 (6). This is similar in design to a single compartment tank except that it is somewhat larger and has a dividing wall within the tank. Slots or ports allow the clear liquid to pass from the first to the second compartment. The principle is that the major portion of the solids will accumulate in the first compartment and only an overflow will be carried into the second compartment. This provides a monitoring site and in general lessens the chances that sludge will be carried into the effluent pipe. When sludge is observed in the last compartment of a multicompartment tank, it is time to have the solids removed from the entire septic tank.

An accessible and removable filter is also available for installation on the effluent baffle (7). This traps fine solids, preventing them from being carried over to the seepage field where

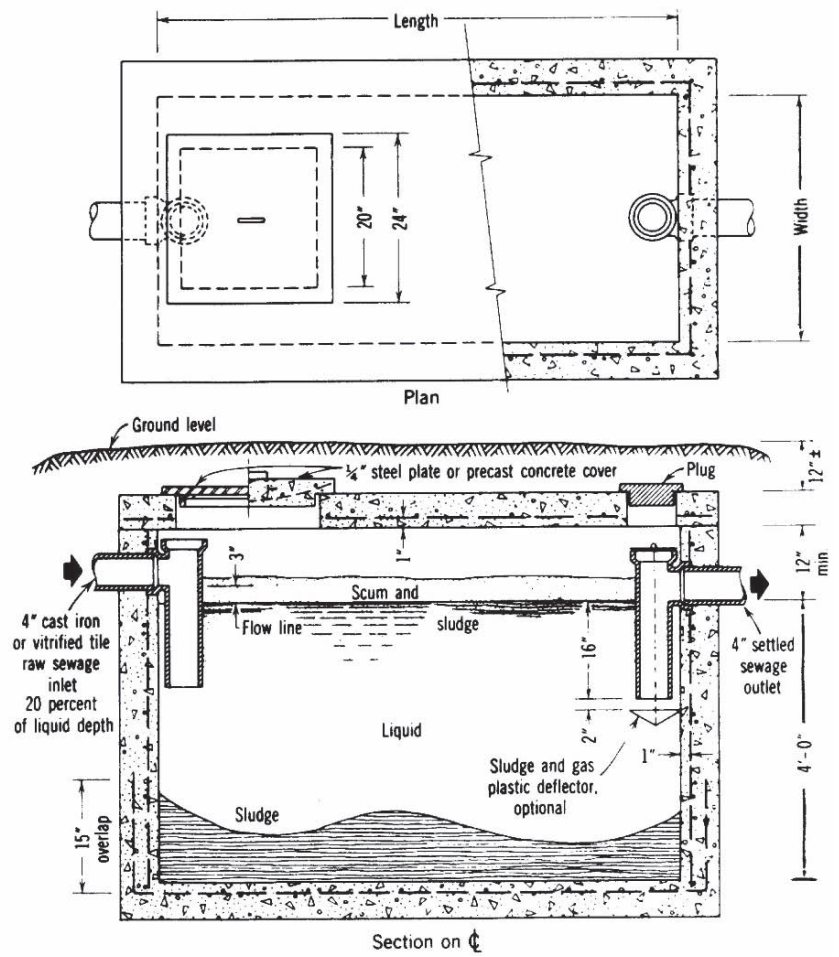


Fig. 3.1. Typical single compartment septic tank (6).

they would cause clogging. Whereas this reduces clogging of the drainfield (which is costly to replace), it somewhat defeats the advantage of a septic tank that requires a minimum of maintenance. Provisions must be made to access the filter, and inspection and cleaning are necessary for it to be effective.

2.1.2. Aerobic Tank

In the aerobic tank, the principles of secondary waste treatment are applied. Under aerobic conditions, bacteria in the wastewater assimilate and metabolize organic compounds for growth, and are settled as a bacterial sludge ("biosolids"). The aerobic tank consists of a well-mixed aeration chamber, followed by a settling chamber. Provision is made for recycling some of the settled bacterial sludge as a "seed" for the aeration chamber. Sufficient air must be introduced to maintain aerobic conditions. As with the septic tank, the aerobic tank must be considered as pretreatment to subsurface disposal of the supernatant liquid effluent. Thus, its

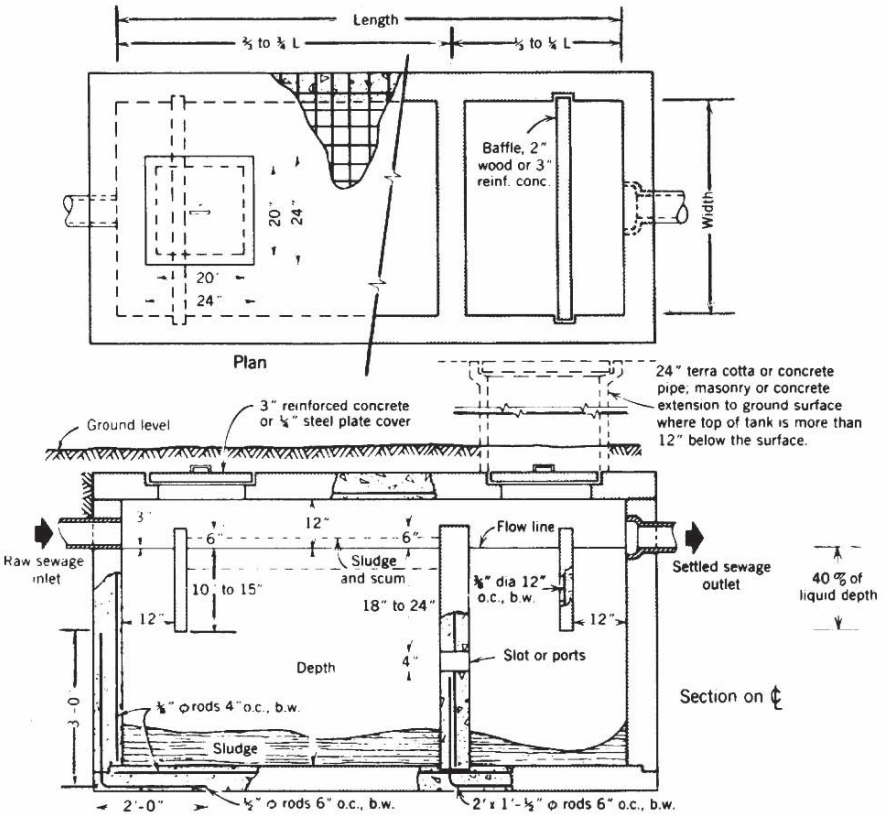


Fig. 3.2. Typical two-compartment septic tank (6).

major function is solids removal. Aerobic treatment affords better solids removal than the septic tank, as is shown in Table 3.1. In addition, significant BOD reduction is obtained. Detention times in the aeration chamber are about 24 hours, with several hours additional time in the settling chamber. An increase in settling chamber detention time can increase solids separation, but at the expense of a decrease in effluent dissolved oxygen. When aerobic treatment was first proposed, sludge buildup was predicted to be negligible. However, some refractory waste components resist degradation, and periodic inspection and sludge removal are necessary (2).

Aerobic tanks are usually factory manufactured and installed in place as needed. Aeration may be provided either mechanically as shown in Figure 3.3 (8), or by means of diffused air as shown in Figure 3.4 (9). The diffused air system in Figure 3.4 is of lightweight fiberglass construction, thereby making it somewhat easier for shipping. There are various other designs of aerobic treatment plants, but the principles are all similar to those shown in the figures.

2.2. *Subsurface Disposal*

The liquid effluent from either septic or aerobic tanks is unsuitable for discharge into surface or ground water, primarily for health considerations. While proposals exist for filtration

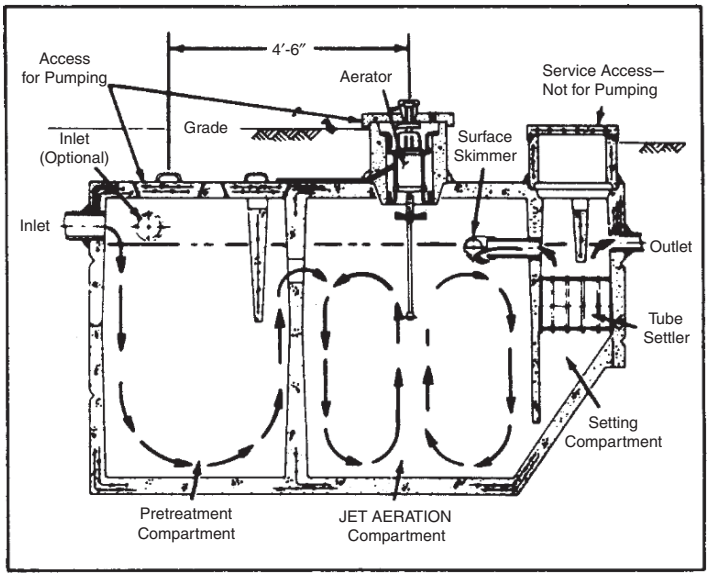
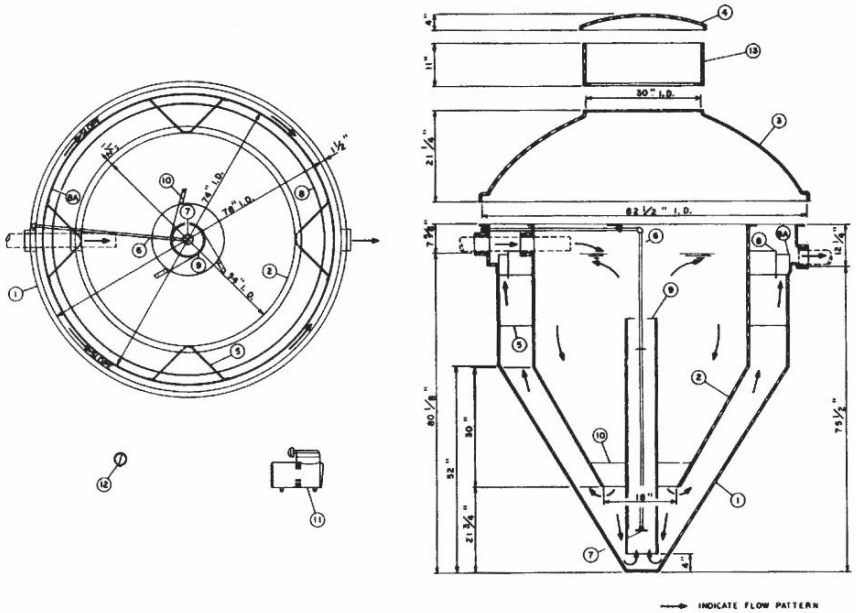


Fig. 3.3. Mechanical aeration treatment plant (7).



PART NO	DISCRIPTION	PART NO	DISCRIPTION
1	OUTER TANK	8A	OVERFLOW WEIR
2	INNER TANK	9	RECIRCULATION PIPE
3	COVER	10	RECIRCULATION PIPE SUPPORT
4	LID	11	COMPRESSOR
5	INNER TANK SUPPORT	12	ALARM PRESSURE GAUGE
6	AIR LINE	13	EXTENSION (OPTIONAL)
7	DIFFUSSER		
8	SCUM BAFFLE		

Fig. 3.4. Aerobic treatment unit (8).

and disinfection of aerobic tank effluent to make it suitable for surface discharge, the possibility of system failure precludes surface disposal. Thus subsurface disposal of the effluent is the only viable alternative. Subsurface disposal represents the final treatment of the waste, and thus must achieve the objectives of the treatment: to prevent contamination of surface and ground water and to avoid nuisance effects.

In all subsurface disposal systems two factors must be considered: the hydraulic loading and the required purification. The hydraulic factor requires the removal of the liquid from the treatment tank through infiltration, percolation, and evapotranspiration. The objective is to prevent emergence of liquid to the surface of the ground. Purification demands a suitable environment for the physical, chemical, and biological processes necessary to remove pathogenic organisms and undesirable chemicals.

2.2.1. Conventional Tile Field

The most common method of subsurface disposal is the tile field, parts of which operate under anaerobic conditions, or at least under limited oxygen availability (Figure 3.5) (6). Several factors are responsible for the hydraulic performance of the drainfield. These factors have been studied extensively by McGauhey, Winneberger, and other researchers (10–14) and include percolative capacity, infiltrative capacity, particle size, and loadings rate.

The percolative capacity is the rate at which water can flow through soil interstices. For years, the percolation test, developed by Henry Ryon in the 1920s, has been the sole determinant of drainfield design. Ryon's percolation test, as described in the *Manual of Septic Tank Practices* (1), is as follows:

1. Six or more tests should be made in holes spaced uniformly over the proposed absorption field.
2. A hole is dug or bored, with horizontal dimensions of 4 to 12 in. (10 to 30 cm), and depth equal to that of the proposed trenches.
3. Sidewalls and bottom of the holes are scratched with a pointed instrument to remove smeared surfaces and to provide a more natural infiltrative surface. Loose material is removed from the hole, and the bottom is covered with 2 in. (5 cm) of coarse sand or fine gravel.
4. The soil is saturated with water to swell the soil to simulate actual conditions of use. The hole is filled at least 12 in. (30 cm) over the gravel, and this level maintained at least 4 hours, and preferably overnight, with an automatic siphon. Percolation rate is determined 24 hours after the first addition of water to the hole, as described in item 5. In sandy soils containing little or no clay, the swelling procedure may be eliminated, and percolation measured as described in item 5c below.
5. Percolation rate measurement
   a. If water remains in hole after the overnight swelling period, the depth of water is adjusted to 6 in. (15 cm) over the gravel. From a fixed reference point, the drop in water level over a 30-minute period is measured. For example, if the level drops 0.75 in. (1.9 cm) in the 30-min period, the percolation rate is calculated:

$$30 \text{ min}/0.75 \text{ in.} = 40 \text{ min /in. drop } (15.7 \text{ min/cm})$$

   b. If no water remains in the hole after the overnight swelling period, the level is restored to 6 in. (15 cm) above the gravel, and the drop is measured at 30-mm intervals for 4 hours, with the

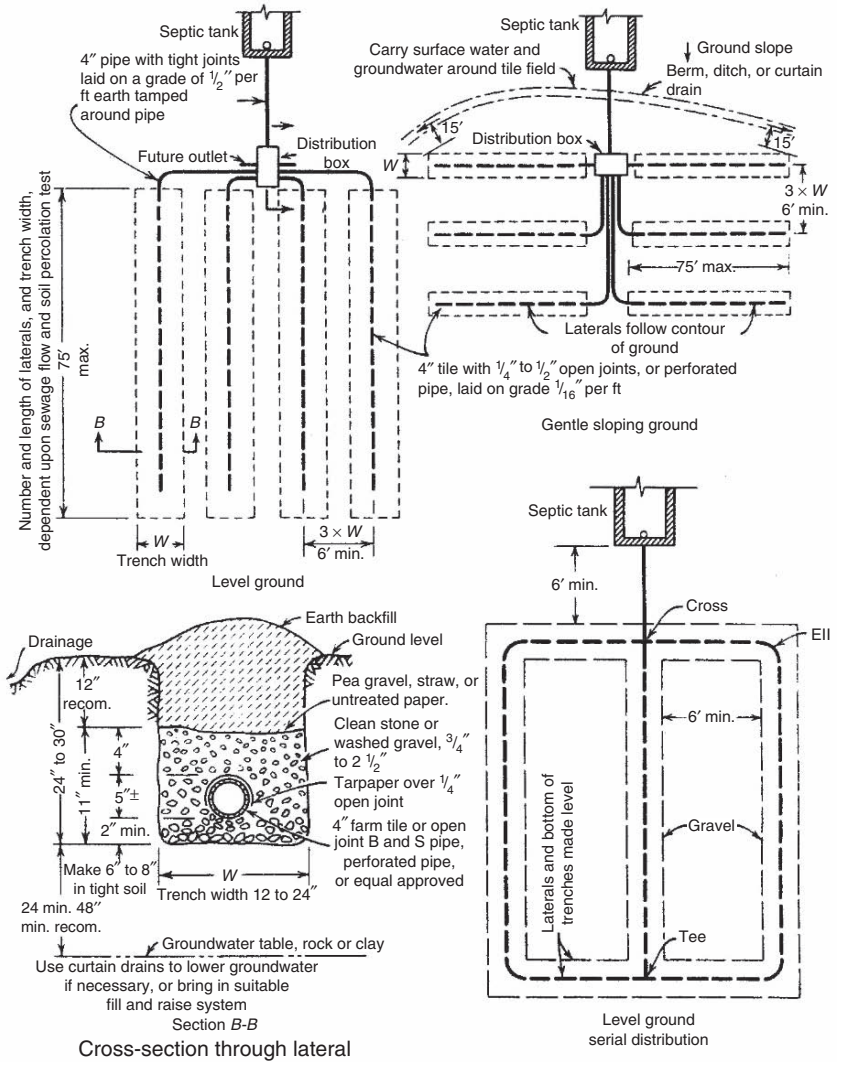


Fig. 3.5. Arrangements and details for tile field disposal systems (6).

level restored to 6 in. (15 cm) above the gravel as necessary. The drop occurring during the final 30-min interval is used to calculate the percolation rate, as in 5a.

c. In sandy soils (or any soil where the first 6 in. of water seeps away in less than 30 mm, after the overnight swelling period), the time interval between measurements is 10 min, and the test is run for 1 hour, with the drop during the final 10-min interval used to calculate the percolation rate. For example, if the drop is 3 in. (7.5 cm) during this time interval, the percolation rate is:

$$10 \text{ min}/3 \text{ in.} = 3.33 \text{ min/in. drop } (1.3 \text{ min/cm})$$

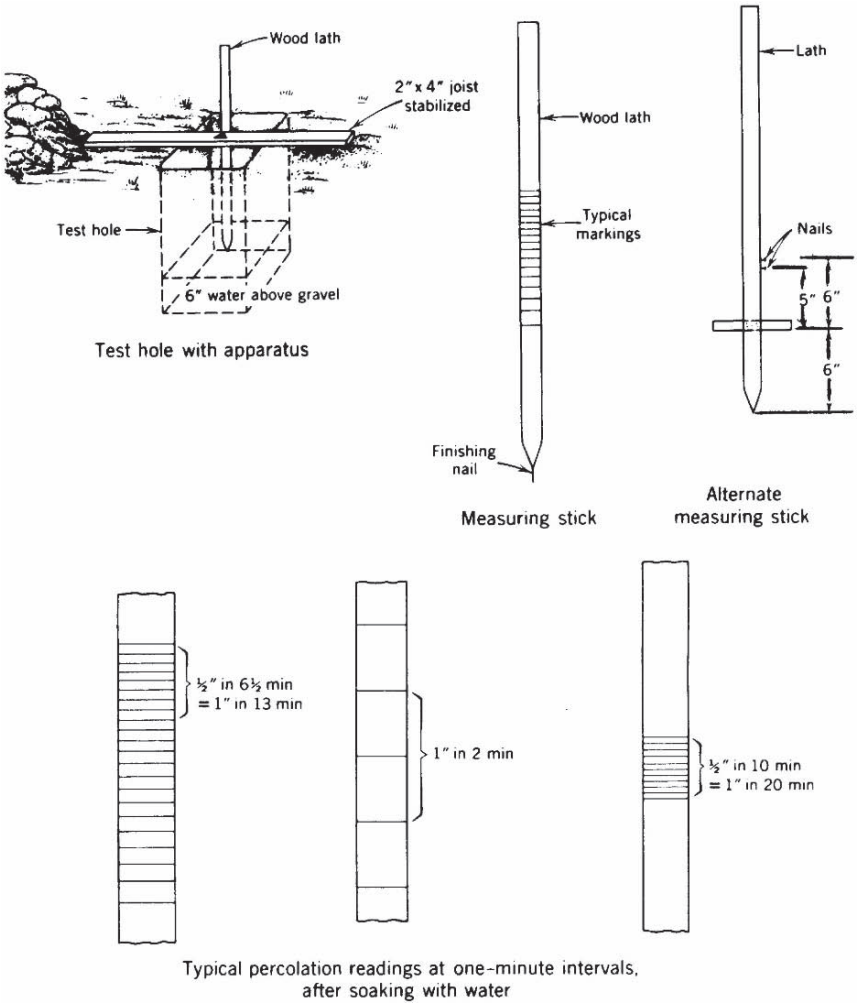


Fig. 3.6. The soil percolation test (6).

A typical installation and some measurement examples are shown in Figure 3.6 (6). The percolation rate is used to calculate the volume of liquid that can be disposed of per unit area of soil, based on the bottom area of the test hole and the drainfield trenches. However, considerable evidence disputes the value of the percolation test as generally applied. Winneberger and Timothy (15) showed that the sidewall area of the test hole and trenches is more important than the bottom area in determining percolation rates, and suggest basing design on sidewall area. Many other factors can affect percolation test results (11, 16). Further standardization of the test procedure is essential.

However, the percolation rate alone is not a sufficient basis for design. The infiltration rate, i.e. the rate of transport of water through the surface on which it is applied, is generally the limiting factor in the hydraulic loading capacity. The infiltration rate can be measured by use

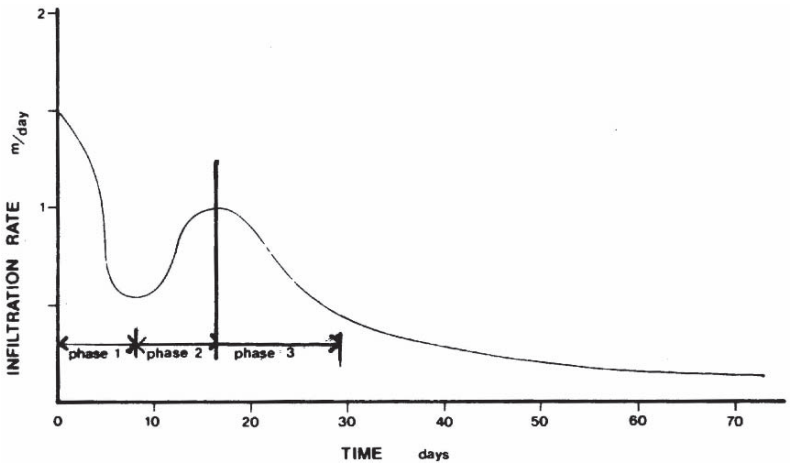


Fig. 3.7. Anaerobic tile field infiltration rate *vs.* time.

of an infiltrometer in which a tube or other boundary isolates a section of soil. Water is applied to the surface, and the infiltration is determined as the difference between the volume applied and the volume remaining after a measured duration of time. The rate is expressed in units of velocity, i.e., cm/h or in/h, calculated from the volume infiltrated per surface area per unit time. For example, if 10 L (2.64 gal) of water is applied to a surface of 1 m^2 (10.8 ft^2), and after a time interval of 1.5 hours, 4 L is measured as remaining or noninfiltrated water; the infiltration rate may be calculated as:

$$(10 - 4)\,\text{L infiltrated}/1\,\text{m}^2\,(1.5\,\text{h}) = 4\,\text{L}/\text{m}^2\,\text{h} = 0.4\,\text{cm/h}.$$

The infiltration rate for soil is not constant, however, but varies with time as a function of the physical, chemical, and biological processes occurring at the effluent-soil interface. Figure 3.7 illustrates these processes. The initial drop in infiltration rate with time (phase 1) is caused by slaking of the soil with water. The following increase in rate (phase 2) is a result of dissolution of trapped gases in the soil, effectively increasing the area available for infiltration. The final long-term decrease in rate (phase 3) is primarily a result of biological activity: the formation of a microbial mat that clogs the surface. Physical processes of compaction and fine particle clogging and chemical processes of deflocculation of soil components and ferrous sulfide production also play a role (11, 17). These clogging effects result in long-term acceptance rates (LTAR) essentially independent of soil characteristics.

The soil particle size, although not directly related to infiltration rate, has an effect on drainfield performance. In fine soils, the organic mat limiting infiltration capacity is rapidly built up at the surface, while in coarser soils it is distributed in depth, slower to build up, and less limiting of infiltration. In addition, in fine soils, capillary forces are increased, thereby requiring a greater distance between the bottom of the drain trenches and the groundwater level to prevent saturation of the drainfield with water.

Finally, the method and frequency of effluent loading on the drainfield can influence performance. Periodic resting of the drainfield allows the establishment of aerobic conditions in the trenches, thus resulting in more rapid decomposition of the clogging mat and restoration of infiltrative capacity. Resting may be achieved by alternate dosing of two drainfields, or through a holding tank with an intermittent dosage system. However, both of these techniques increase the cost and complexity of operation, which are serious drawbacks in the case of individual household treatment systems. In any case, aeration occurs in conventional tile fields because of aeration through the soil or in the distribution pipe. Also, dosing is intermittent because of the normal variable use of water in the home. In general, anaerobic conditions persist only several inches into the original soil in unrested fields.

Biological activity in the drainfield trenches and the surrounding soil is responsible for the final treatment of the effluent. The efficiency of bacterial removal in soil has been studied (18–21), with conclusions that almost total removal occurs within 1 m (3 ft) in aerobic soils. In addition, in fine soils, the microbial mat acts as an effective filter, producing removals in shorter distance, at the price of lower infiltration rates. Sewage bacteria are seldom observed more than 30 m (100 ft) from the point of entry into the soil. The effectiveness of soil in the removal of nutrients (nitrogen and phosphorus) is discussed in a later section.

2.2.2. Aerobic Drain Field

Aerobic conditions in the drainfield have two major advantages: (1) a more rapid stabilization of the effluent and removal of pathogenic bacteria, and (2) higher infiltration rates through decomposition of the microbial mat formed at the soil-water interface. However, maintenance of aerobic conditions in all segments of the tile field can be difficult, particularly if anaerobic tank effluent is applied to it. Bernhart (3) lists the possible sources of oxygen for an aerobic drainfield:

1. Dissolved oxygen in the aerobic tank effluent.
2. Ventilation through wastewater distribution pipes and additional air pipes, all vented.
3. Air supply from the ground surface through the porous seepage bed.
4. Oxygen uptake from air through splashing of tank effluent.
5. Ventilation through distribution pipes during rest periods.

In general, aerobic drainfields are shallow to maximize aeration from the surface, as is shown in Figure 3.8. In a variation of the aerobic bed, the mound system, the drainfield is constructed in sand fill above the natural ground layer to maximize surface aeration. Considerations of hydraulic loading and effluent treatment for this technique are similar in principle to those discussed for the conventional drainfield.

2.2.3. Evapotranspiration Systems (22)

Where risks of groundwater and surface water contamination might exist, evapotranspiration (ET) systems may provide a solution. Effluent from a septic tank or an aerobic treatment unit flows into a vegetation covered distribution and storage area where the liquid is removed by evaporation from the surface of the soil and by transpiration from the plants. ET systems are most effective in areas where annual evapotranspiration exceeds the loading of the system from the combined effluent and rainfall.

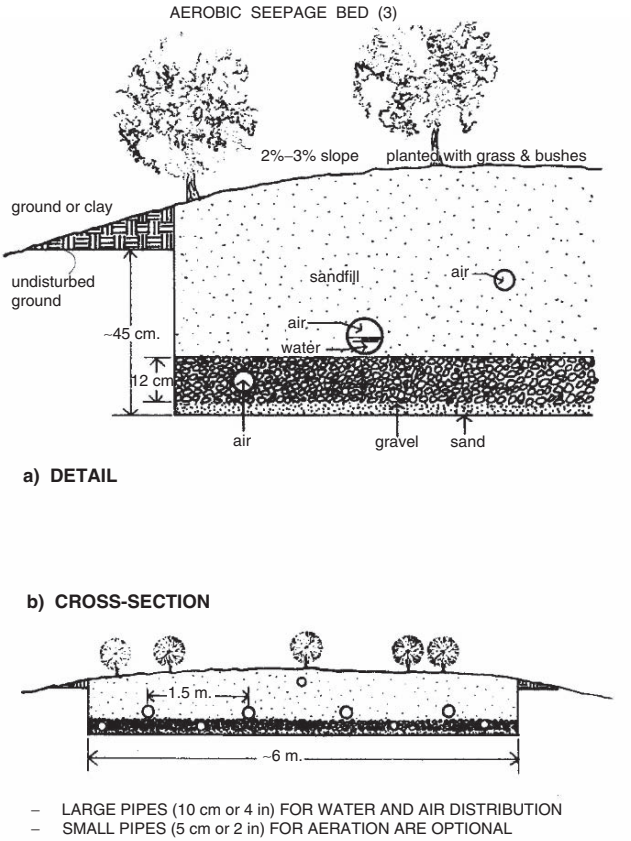


Fig. 3.8. Aerobic seepage bed.

There are two modifications of the ET system. The lined ET system (Figure 3.9) relies strictly on evapotranspiration with no adsorption into the underlying soil. The ETA (evapotranspiration/adsorption) system (Figure 3.10) has no underlying liner, and takes advantage of the additional seepage of the liquid into the ground.

2.2.4. Seepage Pit

The seepage pit, or leaching cesspool (Figure 3.11) (6), is a covered pit with an open-jointed lining through which septic tank effluent can seep into the surrounding porous soil. It should be used only in porous soil where the groundwater table is low, and only if a tile field cannot be provided, e.g., because of insufficient available area. Seepage pits are considered a less desirable disposal method, and are, in fact, banned in some areas (1). Determination of the soil porosity is difficult, and porosity must be determined for each stratum the pit will penetrate. McGauhey and Winneberger studied the failure of seepage pits (11), and concluded that failure due to clogging is inevitable. Clogging proceeds progressively from the bottom of

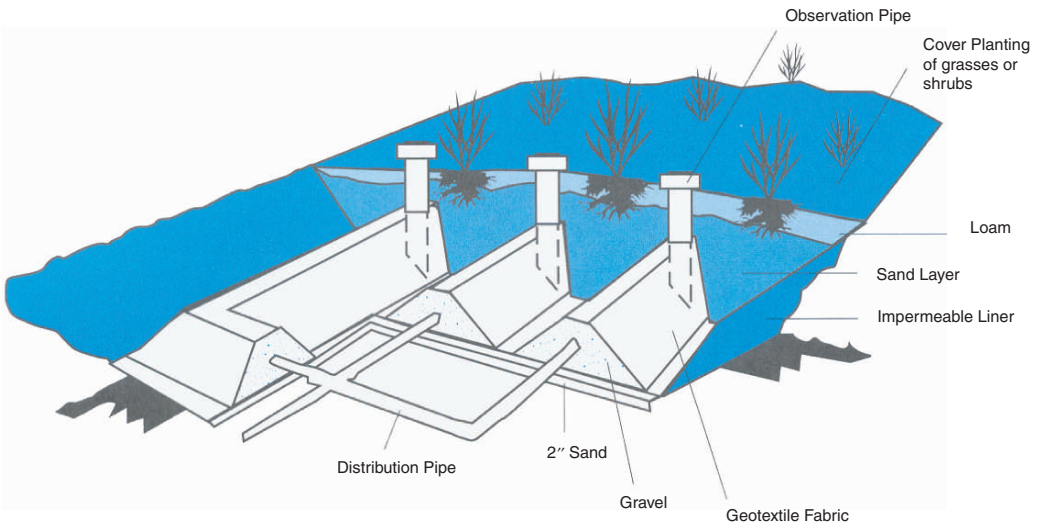


Fig. 3.9. Evapotranspiration (ET) leach field (22).

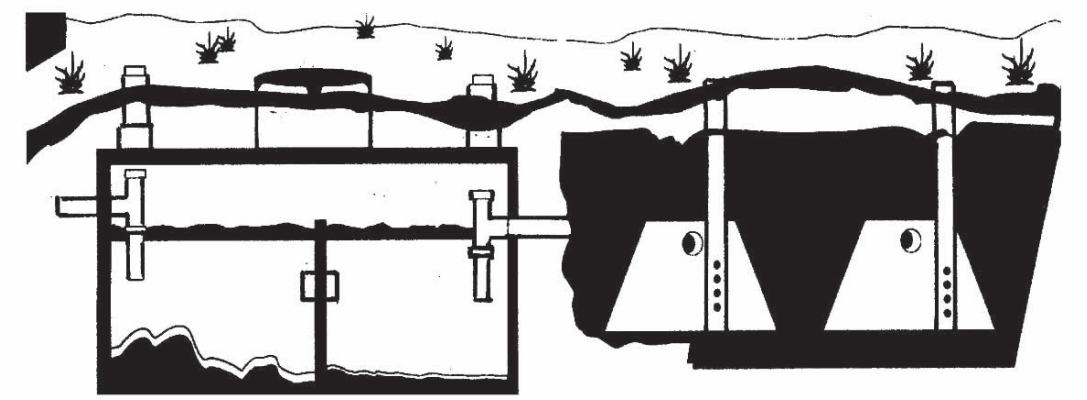


Fig. 3.10. Evapotranspiration/absorption (ETA) leach field (22).

the pit to the top, as lower levels become essentially sealed off from infiltration because of slime buildup. In favorable soils, this process may take many years.

In some areas, leaching cesspools are used without being preceded by a settling tank. In general, these are not recommended except where the soil is extremely porous, or as a temporary expedient. Without the previous removal of the larger solids, cesspools tend to clog rapidly.

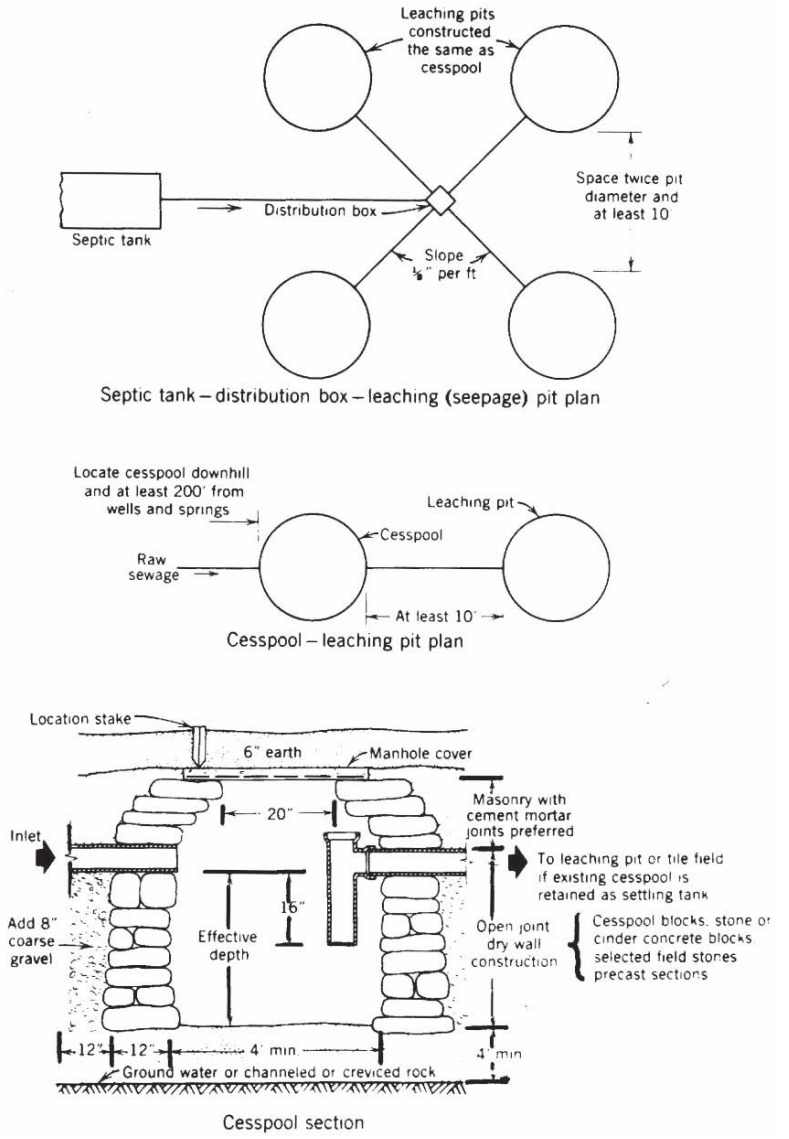


Fig. 3.11. Leaching pit and cesspool details (6).

3. DESIGN

The design practices described in this section are intended to be representative of the basic concepts in the design of an on-site subsurface treatment system. They may vary from those suggested in the *Manual of Septic Tank Practice* (1). The US Environmental Protection Agency (US EPA) has also provided a *Design Manual* (23) and a *Guide to Septage Treatment and Disposal* (24). In addition, the American Society of Civil Engineers (ASCE) has published

a series of *Standards Related to On-site Septic Systems* (25–27). However, state and local regulatory agencies should be consulted particularly if a nonstandard disposal system, e.g., evapotranspiration for effluent disposal, is selected. The National Small Flows Clearinghouse publishes *A Guide to State-Level Onsite Regulations* (28) that provides information on state regulations.

3.1. General Considerations

3.1.1. Flow

The design of an individual wastewater treatment system is based upon the quantity and characteristics of the flow. These may vary widely from household to household, and thus conservative estimates are used for design. Estimation of flow is usually based on the number of bedrooms in the dwelling rather than the number of occupants, because of the likelihood of change in the latter parameter. Average daily per capita water consumption is about 200 L/c-d about (50 gal/c-d), with approximately two persons per bedroom for a two-bedroom dwelling and 1.5 persons per bedroom for each additional bedroom (3). Cotteral and Norris suggest 400 L/c-d (about 100 gal/c-d) as a safe flow estimate (17). Thus design of an individual treatment system can be based on the number of bedrooms in the dwelling.

3.1.2. Site Suitability

Many factors must be considered to insure proper operation of the system. These include lot size, slope, depth to bedrock, depth to groundwater table, and location of wells, water bodies, and other structures.

3.1.2.1. Lot Size

Sufficient lot size must be provided to insure adequate suitable drainfield area and the necessary distance between septic tank and seepage field and wells and structures (see Section 3.1.2.5). In addition, wherever possible, suitable area for a replacement or exchange drainfield should be reserved. In general, a lot size of about 0.4 ha (1 ac) is sufficient to satisfy these requirements.

3.1.2.2. Slope

Because of construction difficulties on slopes and to insure proper effluent distribution, additional drainfield area should be reserved wherever the ground surface slope exceeds 5%. Also, wherever possible, the treatment system should be downslope of wells in the area.

3.1.2.3. Depth to Bedrock

Rock formations or other impervious strata must be at a depth greater than 1.3 m (4 ft) below the bottom of tile field trenches or seepage pits (1).

3.1.2.4. Depth to Groundwater Table

The seasonal high groundwater table must be at least I m (3 ft) below the bottom of tile field trenches, and 1.3 m (4 ft) below the bottom of seepage pits. This is particularly important in fine soils where capillary action may cause saturation of the field.

Table 3.2
Setback requirements (3)

	Federal Housing Authority		Uniform Plumbing Code		Proposed Marin Co., CA, standards (17)	
	m	ft	m	ft	m	ft
A. From Septic Tanks						
Buildings	1.5	5	1.5	5	1.5	5
Property lines	3	10	1.5	5	1.5	5
Wells	15	50	15	50	30	100
Creeks, streams	—	—	15	50	1.5	5
Cut, embankments	—	—	—	—	7.5	25
Pools	—	—	—	—	3	10
Water lines	3	10	1.5	5	3	10
Walks, drives	—	—	—	—	1.5	5
Large trees	—	—	3	10	3	10
B. From Drainfields						
Buildings	1.5	5	2.5	8	3	10
Property lines	1.5	5	1.5	5	1.5	5
Wells	30	100	15	50	30	100
Creeks, streams	—	—	15	50	30	100
Cut, embankments	—	—	—	—	30	100
Pools	—	—	—	—	7.5	25
Water lines	3	10	1.5	5	3	10
Walks, drives	—	—	—	—	1.5	5
Large trees	—	—	3	10	3	10

3.1.2.5. Setback Requirements

To assure that the public is protected from possible effects of pathogenic bacteria, minimum setbacks of septic tanks and drainfields from water supplies and other structures have been established. Table 3.2 contains setback requirements (3) and those proposed for Marin County, California (17). Particular attention is given to wells, reservoirs, and possible points of entry of effluents into watersheds.

3.2. Septic Tank Design

The major consideration in septic tank design is capacity. The tank must provide sufficient detention time for settling and digestion of solids, as well as capacity for sludge and scum storage (Figure 3.1). Based on a design flow of 400 L/c-d (about 100 gal/c-d), suggested tank hydraulic detention times range from about 1 to 2 d, with an additional 35% to 50% of this volume for sludge and scum storage. Current design practice favors larger tanks. While installation cost is not much greater than for smaller tanks, the increased capacity results in (a) longer detention time; (b) lower suspended solids in the effluent, resulting in less clogging of the tile field; (c) greater sludge and scum storage, requiring less frequent

Table 3.3
Septic tank capacities (16)

	Federal Housing Authority	US Public Health Service (1)	Uniform Plumbing Code	Marin Co., CA, standards
Minimum size, gal				
1–2 Bedrooms	750	750	750	1200
3 Bedrooms	900	900	1000	1200
4 Bedrooms	1000	1000	1250	1200
5 Bedrooms	1250	1250	1500	1450
Additional bedroom	250	250	150	250
Material	Durable	Concrete, metal	Durable	Concrete

pumping; and (d) in creased resistance to upset because of shock loadings. Separation of the tank into two or three chambers (Figure 3.2) is advisable to insure that the effluent does not contain sludge resuspended by "boiling" (gas production from rapid digestion). Tanks are usually constructed of coated steel or concrete. Many precast concrete tanks are available on the market, although on-site construction is sometimes used. Details of tank construction can be obtained in the *Manual of Septic Tank Practice* (1), most state and many local health departments. Septic tank capacities designed on the basis of number of bedrooms are shown in Table 3.3. These capacities allow for all household appliances including garbage grinders.

Capacities of tanks are listed in gallons, because marketed units are sold in terms of this unit. To obtain capacity in liters, multiply number of gallons by 3.8.

3.3. Aerobic Tank Design

Design of an aerobic tank is more complex. Not only must sufficient detention time for stabilization of the waste be provided, but also an oxygen supply system to maintain aerobic conditions and a sludge recycling device are required. Because of the more rapid rate of decomposition under aerobic conditions, tank volume may be about one-half that of an anaerobic unit. Bernhart (3) discusses the details of aerobic tank design. In general, based on a design flow of 400 L/c-d (about 100 gal/c-d) about 1 day total detention time must be provided. The aerobic unit (Figure 3.12) is divided, consisting of a well-mixed aeration chamber(s) with a detention time of 1/2 to 3/4 days, followed by a quiescent settling chamber with 2 to 6 hour detention time. The possible occurrence of instability in the aerobic process leading to insufficient treatment of the effluent requires that a fail-safe device be installed to protect the drainfield. In general, a small gravel or crushed stone filter compartment having about 1.5 hour detention time serves this purpose. Pretreatment of the wastewater by settling before aeration is optional.

Oxygen is supplied in the aeration chamber by either mechanical mixing (Figure 3.3) or compressed air diffusers (Figure 3.4). About 150 m^3 of air is required for each kg BOD removed (about 2400 ft^3/lb BOD). Each person contributes about 0.12 kg of BOD/d

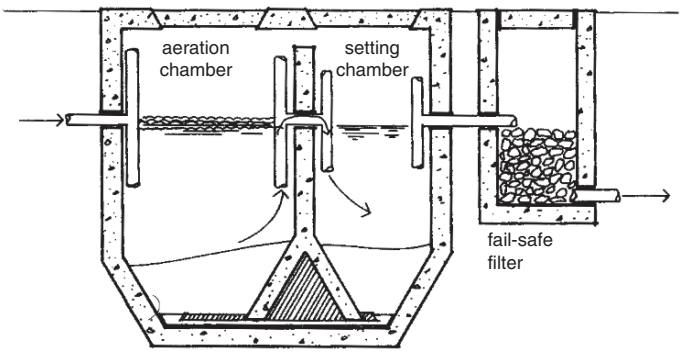


Fig. 3.12. Aerobic tank.

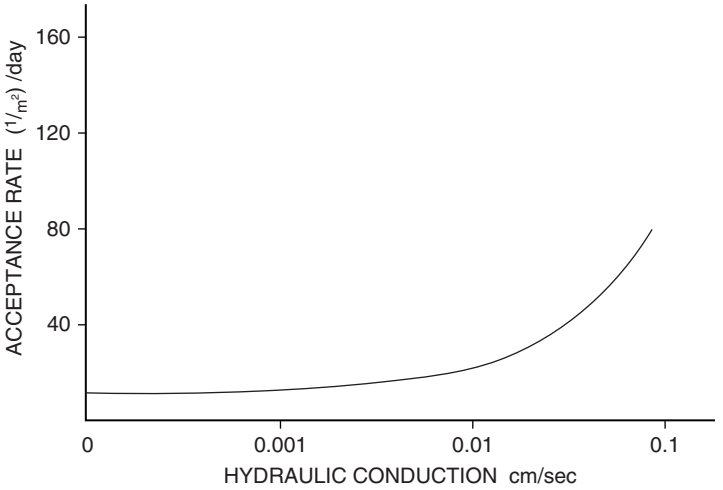


Fig. 3.13. Anaerobic tile field long-term acceptance rate *vs.* soil hydraulic conductivity (29).

(0.3 lb BOD/d). Sludge recycle is accomplished either by gravity feed or pumping. Some systems allow regulation of sludge recycle. Some units provide for posttreatment of the effluent by disinfection; however, this seems inadvisable, because underground disposal is still required, and disinfection may destroy the biota necessary for further treatment of the effluent. Many aerobic treatment units are available on the market.

3.4. Conventional Tile Field

The conventional anaerobic tile field has been the most common source of failure in individual treatment systems. Studies indicate that design based on the percolation test and the bottom area of the trench is incorrect. In operation, the long-term infiltration rate and the trench sidewall area available for infiltration are the more important parameters.

Table 3.4
Drainfield design based on *Manual of Septic Tank Practice* (1)

A. Trench bottom area per bedroom

Percolation rate		Trench bottom area per bedroom	
min/in.	min/cm	ft^2	m^2
1 or less	0.4 or less	70	6.51
2	0.8	85	7.90
3	1.2	100	9.3
4	1.6	115	10.69
5	2.0	125	11.62
10	3.9	165	15.33
15	5.9	190	17.66
30	11.8	250	23.23
45	17.7	300	27.88
60	23.6	330	30.67
Over 60	Over 24	Not suitable	

B. Required trench spacing

Trench width		Minimum distance between trench center lines	
in	m	ft	m
12–18	0.3–0.5	6.0	1.83
18–24	0.5–0.6	6.5	1.98
24–30	0.6–0.75	7.0	2.13
30–36	0.75–0.9	7.5	2.29

As discussed in the theory section (Section 2), the formation of a biological mat leads to long-term infiltration rates almost independent of percolation rate. Healy and Laak (29) have correlated the results of many studies of long-term infiltration in various soil types and found rates ranging from 12 $L/d/m^2$ (0.3 gpd/ft^2) for clay loam to 32 $L/d/m^2$ (0.8 gpd/ft^2) for fine sand (see Figure 3.13). Thus for soils varying in hydraulic conductivity by a factor of 100, the long-term infiltration rate varied only by a factor of three. Coarser sands showed higher rates. These results can be used for design of a tile field that should operate indefinitely.

Despite its limitations, drainfield design based on Ryon's percolation test is in widespread use. Using this test, the drainfield area is determined from the required bottom area of trenches per bedroom, the trench spacing, and the trench width. Table 3.4 lists the required trench bottom area per bedroom for various percolation rates, and the required trench spacing. Minimum design is for two bedrooms, regardless of actual circumstances. Depth of trenches is at least 18 in. (0.5 m) below the surface, with at least 6 in. (15 cm) of gravel below the drain tile, and 2 in. (5 cm) above.

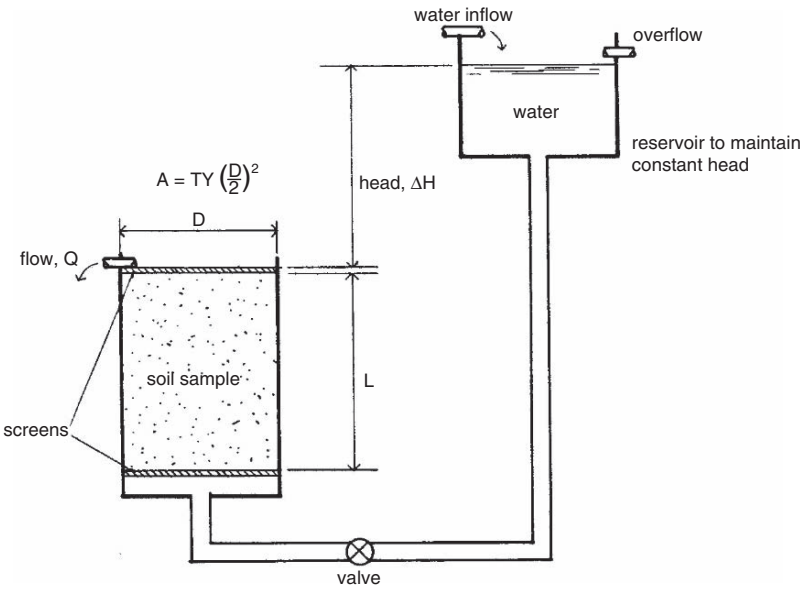


Fig. 3.14. Upflow permeameter.

The suggested design approach is to test the soil permeability using the conventional percolation test or a standardized hydraulic conductivity test. Hydraulic conductivity can be measured in a permeameter (Figure 3.14), essentially a tube in which the soil sample is subjected to a constant hydraulic head of water, and the resultant flow through the sample is measured, with units of flow/area of cross section, e.g., m^3/m^2-d or gpd/ft^2. The test is rapid; however, it is difficult to assure that the sample is representative of the area in question, and some changes (e.g., compaction) may occur on introduction of the sample to the tube. On a large scale, permeability may be measured in the field by dye or tracer studies, or by pumping tests in wells.

If a permeameter is used to determine the hydraulic conductivity of a soil sample, referring to Figure 3.13, the hydraulic conductivity, K is calculated from the relation:

$$K = \frac{Q}{AS} = \frac{QL}{A\Delta H} \tag{1}$$

where Q is the volumetric flow rate through the sample cell (m^3/d or gpd), A is the cross-sectional area of the cell, (m^2 or ft^2), S is the hydraulic slope, L is the cell height (m or ft), and H is the hydraulic head (m or ft). As an example, if the permeameter cell has a height of 0.1 m, a cross-sectional area of $0.0025\,m^2$, and an applied hydraulic head of 1 m water, and a flow of $1\,m^3/d$ is measured for a given soil sample, the hydraulic conductivity of this sample is:

$$K = QL/A\,\Delta H = (1\,m^3/d)(0.1\,m)/(0.0025\,m^2)(1\,m) = 40\,m^3/m^2\text{-d}\,(977\,gpd/ft^2)$$

Table 3.5
Anaerobic tile field design (17)

	US Public Health Service (1) FHA[a]	Uniform Plumbing Code[a]	Marin County. CA standard[b]	Most conservative approach[c]
Percolation test used	Yes	As required	As guide	As guide
Design surface	Bottom	Bottom	Sidewall	Sidewall
Trench width, cm	30 to 100	30 to 100	30 to 45	30 to 45
Gavel depth below tile, cm	15	30	70 to 100	70 to 100
Minimum sidewall area, m^2	13 (140 ft^2)	18 (200 ft^2)	(1700 ft^2)	—
Sidewall area per bedroom, m^2	18 to 30	7 to as required	55	75, 55[d]
Minimum trench spacing, m	2	2	2	2
Distribution box used	Yes	Yes	No	No

[a] Design based on percolation rate and bottom area. Sidewall areas calculated on basis of minimum depth and width. Variance in sidewall area is for percolation rate range of 15 to 60 min/in.

[b] Design approximated on design flow of 400 Lpcd with occupancy equal to number of bedrooms plus one, and safe loading rate of 11 $L/d/m^2$ for sidewall area. This may be considered a moderately conservative approach.

[c] Conservative approach based on design flow of 400 Lpcd, wit occupancy of 2 persons in each of first 2 bedrooms, 1.5 in each additional bedroom. Safe loading rate of 11 $L/d/m^2$ is used.

[d] 75 m^2 for each of first 2 bedrooms, 55 m^2 for each additional bedroom.

The permeameter test does not allow for sidewall infiltration. Field observed percolation rates of more than about 50 min/cm (120 min/in.) or hydraulic conductivities of less than 5×10^{-5} cm/s (10^{-4} ft/min) indicate Insufficient permeability for a tile field. With the hydraulic conductivity for a particular site established, a conservative wastewater loading design factor of 11 $L/d/m^2$ (0.25 gpd/ft^2) may be used as the long-term acceptance rate for any soil. The infiltrative area is the trench sidewall area below the backfill. If one wishes to take advantage of the higher long-term acceptance rates applicable to more permeable soils (Figure 3.13), the permeameter test (see Figure 3.14) must be employed, and a safety factor of about 25% should be incorporated into the design infiltration rate.

Healy and Laak (29) have also considered the possibility of the hydraulic conductivity of the soil acting to limit infiltration. This may occur in areas of high water table and very low hydraulic conductivity. If a soil's hydraulic conductivity is less than 10^{-4} cm/s (2×10^{-4} ft/min) and the distance from the bottom of the trench to the high water table is less than 1.6 m (5 ft), this effect may occur. Drain field design in areas of low permeability and high water table should consider this effect.

To provide sufficient sidewall area for infiltration, some modification of standard practice is necessary (see Table 3.5). The depth of rock fill below the drain tiles is increased to increase the effective sidewall area, and trenches are narrower to increase the surface-to-volume ratio. The bottom of the trenches must remain 1 m (3 ft) above high groundwater table. The practical limit for trench depth below the surface is about 2 m (6 ft). In general, 4 in. drain tile is used, though numerous varieties of tiles and perforated pipes are available. The bottom of the trench may be covered with light gravel, with rock fill extending 0.6 to 1 m (2 to 3 ft) below the tile and at least 5 cm (2 in.) above. The top of the rockfill is covered with paper or straw to prevent

silt from filling the trench, and 0.3 to 0.5 m (12 to 18 in.) of backfill covers the trench. Trench units are usually 15 to 30 m (50 to 100 ft) in length, with a slope of about 0.2%. Parallel trenches must be at least 2 m (6 ft) apart and should be separated by twice the depth for deep trenches.

Accepted practice once favored parallel distribution of flow to sections of trench through the use of a distribution box. However, such a system may force flow to a failed trench section, thereby causing breakthrough of effluent to the surface. Series distribution of flow is now recommended, where the full capacity of one trench section is utilized before the liquid overflows into the next section. Figure 3.15A illustrates the recommended flow distribution on essentially level ground, with modifications for sloping ground shown in Figure 3.15B. Note that trenches run across the slope, with upper sections filled before overflow to lower sections occurs. Successive trench sections in line should be separated by undisturbed ground a distance of twice the depth. All trench sections as well as the line from the septic tank to the trenches should be connected with watertight pipe. With the noted exceptions of trench depth and series distribution, the construction guidelines in the *Manual of Septic Tank Practice* (1) should be followed. Provision must be made for diversion and drainage of storm water from the active infiltration area.

The above design is based on use of septic tank effluent. No definite work has discussed the allowable design modification of the anaerobic tile field if loaded with aerobic tank effluent, although it is generally accepted that the lowered suspended solids and BOD loadings lessen field clogging. Data in Bernhart (3) suggest that drainfield size may be decreased by one-half if aerobic effluent is used, and this work should be consulted if such a system is desired. It is more common that an aerobic tank be followed by an aerobic drainfield to utilize fully the advantages of aerobic treatment.

3.5. Aerobic Tile Field

The aerobic tile field generally takes the form of a crowned seepage bed, although trench systems are possible. In the aerobic seepage bed (Figure 3.8), the bed is excavated 0.3 to 0.5 m (12 to 18 in.) below original ground level. The floor is level, and is covered with 3 to 5 cm (1 to 2 in.) of sand and 15 cm (6 in.) of gravel or crushed stone. Water distribution pipes are 4-in. perforated pipe, and are vented to allow air distribution above the wastewater level in the pipe. Additional air distribution pipes may be used. Water distribution pipes must be sufficient to distribute the effluent across the bed; in general, parallel pipes 1.5 m (5 ft) apart the length of the field will suffice. The pipes are covered to 10 cm (4 in.) above the original ground level with sand. The bed is crowned with sand to insure a surface slope of 2% to 3% and covered with up to 5 cm (2 in.) topsoil. The surface is planted with grass and low bushes for ground cover as well as to increase evapotranspiration. If the bed is to be loaded with septic tank effluent, extra provision for aeration must be considered. Batch loading to allow rest and reaeration of the bed is advisable in this case.

In the aerobic seepage bed, only the bottom area is available for infiltration, and thus this area is the major design parameter. Bernhart (3) bases design on the area calculation:

$$A = QP/s \tag{2}$$

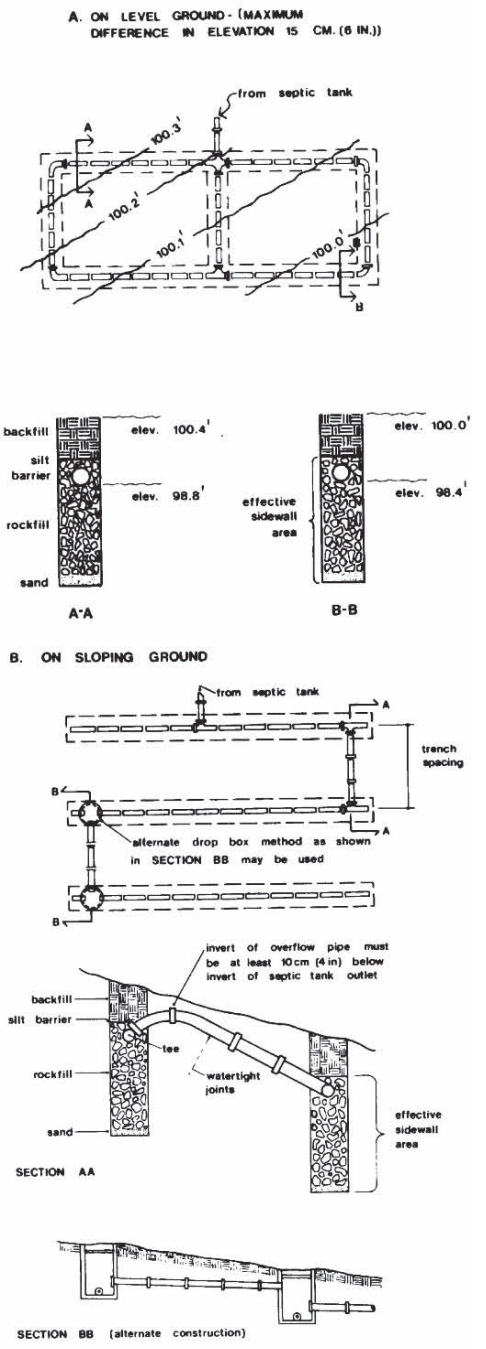


Fig. 3.15. Anaerobic tile field serial distribution layout.

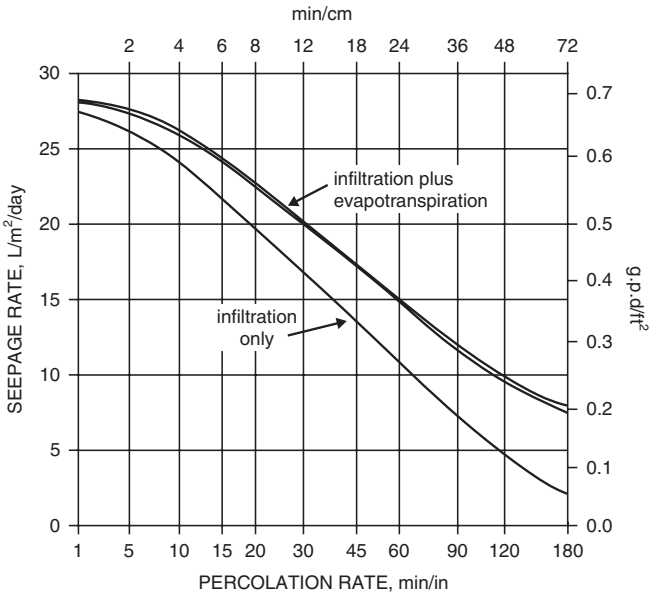


Fig. 3.16. Seepage rates for aerobic beds (3).

where A is area (m^2 or ft^2), Q is the wastewater flow rate (L/d or gpd), P is the tank effluent pollution load, and $s$is the seepage rate ($L/d/m^2$ or gpd/ft^2). The pollution load, P, is an empirical dimensionless parameter:

$$P = (\mathrm{BOD} + \mathrm{SS})/120 \tag{3}$$

where BOD and SS are biochemical oxygen demand and suspended solids, respectively, both in mg/L. For a properly operating aeration tank, P ranges from 0.9 to 1.2; for well-designed septic tanks from 1.8 to 2.8; and for small septic tanks from 3.0 to 3.8. The seepage rate, s, is the sum of infiltration and evapotranspiration. Values of s versus percolation rate (6 in. or 15 cm hole) for Toronto, Ontario, Canada are shown in Figure 3.16. The values are representative of the northern United States. Bernhart (3) lists factors for calculation of the evapotranspiration rates in other areas. The rates shown in Figure 3.16 can be considered conservative. Breakthrough of effluent to the surface should not occur, even with the lowered evapotranspiration of a Canadian winter, although the liquid level may be considerably higher in winter, and the soil may be spongy.

From the above relations, the required seepage bed area can be determined. If the area is larger than is convenient (greater than about 140 m^2 or 1500 ft^2), two or more beds connected in series may be used, preferably with provision for parallel operation. On slopes, terraced arrangements are suitable. Shallow trenches otherwise similar to the anaerobic tile field may be used to provide aerobic drainage. These trenches may be about 0.5 m wide, 0.6 to 1 m deep, and about 1.5 m apart (18 in. wide, 2 to 3 ft deep and 5 ft apart). Infiltration area may be based on sidewall as well as bottom area. This will, however, require more surface area

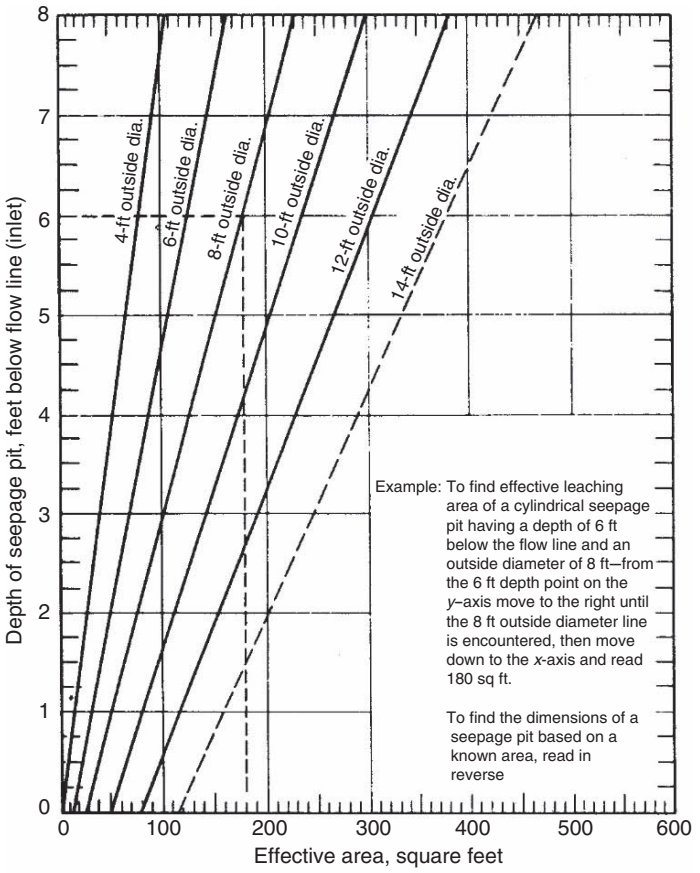


Fig. 3.17. Effective areas of round seepage pits preceded by septic tanks of cesspools (6).

than the seepage bed. Bernhart (3) suggests a trench system around the property perimeter supporting a hedge as a practical aerobic trench system with good evapotranspiration. Further details for design and construction of aerobic seepage beds and trench systems are available in his text (3).

3.6. Seepage Pit

Seepage pits (Figure 3.11), though considered the least suitable method of subsurface disposal of tank effluent, are allowed in many areas. The *Manual of Septic Tank Practice* (1) includes details of construction. The major design parameter is the outer sidewall area of the pit, which may be determined using Figure 3.17. The sidewall area required is determined by a weighted average of the percolation rate for each vertical stratum to be penetrated. No allowance is given for strata of percolation rates slower than 12 min/cm (30 min/in.), and the average must be 12 min/cm or greater or the site is considered not suitable. Sidewall area required per bedroom is shown in Table 3.6. Sizes of pits range from 1 to 4 m (3 to 13 ft)

Table 3.6
Seepage pit design (1)

Percolation rate		Sidewall area required per bedroom	
min/cm	min/in.	m^2	ft^2
0.4 or less	1 or less	6.5	70
0.8	2	8	85
1.2	3	9.2	100
1.6	4	10.6	115
2	5	11.6	125
4	10	15	165
6	15	17.5	190
12	30	23	250
Greater than 12	Greater than 30	Not suitable	Not suitable

diameter and up to 6 m (20 ft) in depth. The bottom of the pit must be at least 1.3 m (4 ft) above the high groundwater table. If one pit cannot provide sufficient area, two or more in parallel may be used, separated by three times the pit diameter, or at least 6 m (20 ft) for deep pits. Seepage pits cannot be used if groundwater contamination is likely, and should not be used if other systems are feasible. State and local authorities should be consulted.

Design of a cesspool without tank pretreatment is similar.

3.7. Institutional and Multiple Dwelling Systems

Although it is beyond the scope of this chapter to include full design criteria, some consideration must be made of the problem of waste disposal from larger housing developments that are not sewered. Required treatment may range from a simple enlargement of the previously described systems to a small sewage treatment facility. The most important design criterion is the waste flow rate. Estimates of flows for most cases can be obtained in the *Manual of Septic Tank Practice* (1). Wastewater characteristics may also be important. A system that is to receive considerable kitchen waste should include a grease trap at the source of the waste.

Sewer lines are necessary to connect all waste sources to the treatment system. Design of the septic tank is similar in principle to that for individual dwellings. The site may play an important role in design, particularly because the large flows entailed may involve a large area for subsurface disposal. A dosing tank may assist in maintaining aerobic conditions in the drainfield by providing rest periods. Where insufficient drainfield area is available, underdrained sand filters may be utilized as final treatment, with chlorination of the effluent from the filter bed. The *Manual of Septic Tank Practice* (1) includes details for construction of these systems.

In some cases, other forms of treatment may be favored. The lower area requirement and more complete treatment afforded by aerobic systems suggest their use. The larger the system, the greater the flexibility of design, because economy of operation becomes more important than the need for a maintenance-free system. For a large installation, any suitable treatment

method should be considered. Some are discussed below. State and local authorities should be consulted.

3.8. Construction

General construction considerations may be obtained from the *Manual of Septic Tank Practice* (1), the US EPA *Design Manual* (23), the US EPA *Guide* (24), the ASCE *Standards* (25–27), or state and local authorities as outlined in the National Small Flows Clearinghouse *Guide* (28). For aerobic treatment systems, further details may be obtained in Bernhardt (3) or from manufacturers.

4. STATE OF THE ART

Research continues on alternatives to conventional septic tank treatment systems. The development of aerobic treatment processes represents one result of this research. Current research considers the problems of improving tank treatment and settling, providing for effluent disposal where conditions preclude conventional drainage fields and preventing pollution of surface and ground water from nutrients contained in the effluent.

4.1. Tank Treatment

The septic tank has changed little since its development. Improvements in treatment have been made through improved baffling and separation of the tank into sections to prevent short-circuiting and improve conditions for settling. Providing larger tanks results in improved tank treatment. Another modification of septic tank treatment includes a compartmented tank that separates sanitary and kitchen wastes from wash wastes ("gray water"), thereby providing much longer detention time for treating the sanitary wastes. After concentrated biological treatment, the sanitary waste is mixed with the wash wastes in another compartment, and treatment is completed (2).

The primary treatment process in a septic tank is the separation of solids and floatables from the liquid portion. Some biological breakdown of the solids occurs. Thus, discharge of nondegradable materials or slow to degrade materials should be avoided. Similarly, large amounts of oils or grease should not be discharged. Ground garbage may be acceptable if sufficient tank capacity is provided, but more frequent pumping may be required. Additives have not been proven to be helpful. Additives that dissolve sludge and/or grease may just carry them over to the leach field, resulting in clogging of the more expensive to repair portion of the system. Starter enzymes have little effect, as the biological content of human wastes is adequate to initiate the biological breakdown of the solids. Oppositely, substances to reduce discharge of pathogens may also reduce the desired biological effect of sludge breakdown in the tank and biological purification in the leach filed mat. Strong household cleaners should be avoided. Here, again, a larger tank will dilute the impact of the discharge of any undesirable substance to the system. Water saving devices in the household will reduce the amount of liquid that must be absorbed by the leach field.

Aerobic tank treatment is subject to considerable modification in design. At present, mechanical mixing or air diffusion are usually used in the aeration chamber. Some current

designs are based on the rotating biological contactor principle, in which a partially submerged medium is rotated through the waste such that during part of its cycle the surface is in contact with air for aeration. The surface of the filter is the active site for biological treatment (2, 3). In a number of aerobic tanks, tubes or plates have been added in the settling chamber to improve sedimentation.

4.2. Effluent Disposal

The problem of effluent disposal in unfavorable areas has received much consideration. Bouma (30) has reviewed the problems associated with shallow soil layer above bedrock, high ground water table, and hydraulics of various soil types, and proposes modifications to overcome these problems. Disposal of effluent in impermeable soils may be achieved through use of artificial sand beds, evapotranspiration of all effluent, or surface disposal of the effluent. Where underdrained sand filters are used, chlorination may be necessary if the filter effluent is to be discharged into a water supply or recreation area. Hines and Favreau (31) discuss a modification of sand filtration for individual home use that employs recirculation of filter effluent to attain desired quality.

Bernhart (3, 32) discusses the feasibility of total evapotranspiration of wastewater, including design parameters. A package system on the market guarantees total evapotranspiration in impermeable soils (2). Surface disposal may take the form of spray irrigation of septic tank effluent in rural areas (33). Aerobic lagoons (33, 34) may be used either as final treatment, or as pretreatment before soil disposal. Some systems make use of the high quality effluent available from extended aeration, and provide chlorination to produce an effluent suitable for surface discharge (35).

Natural systems, such as wetlands, have been shown to reduce BOD, lower pathogenic organisms, and absorb nutrients (36). Wetlands may be either natural or constructed. Proper design must be followed to assure that the waterbody receiving the final effluent is not impaired.

4.3. Nutrient Removal

The fate of the nutrients, primarily nitrogen and phosphorus, present in the effluent from individual household systems has become a matter of considerable concern because of the association of these elements with the accelerated eutrophication of surface waters. Nitrate contamination of wells may also be a serious problem. The movement of septic tank effluent nutrients through soil and groundwater (37, 38) and treated sewage components through sand beds (39, 40) has been studied. Effluent nitrogen from anaerobic treatment is in the form of ammonia and organic nitrogen, with little nitrate present. Some denitrification may occur, releasing nitrogen gas under anaerobic soil conditions. Most of the nitrogen, however, is converted to nitrate under aerobic conditions in the surrounding soil, and in this form may travel considerable distance, with essentially no uptake by the soil or rock through which it flows. In some areas of high ground water or low permeability soils, ammonia may show considerable mobility. Most nitrogen in an aerobic tank effluent already exists in the highly mobile nitrate form.

Most effluent phosphorus is biologically converted to soluble phosphates upon passage through soil systems. However, in general, most soils show a high affinity for phosphates (37–42), and in many cases uptake by soil of effluent phosphates prevents phosphorus contamination of ground and surface waters. However, under conditions of high water table, coarse sand and gravel subsoil, or long-term loading, phosphorus pollution may occur (37). Such conditions may be particularly significant in the case of disposal systems for cottages on lake shores, where proximity to the water body and a high water table may cause transport of considerable phosphorus to the body of water before it has had a chance to be removed by the soil.

Preliminary investigations have been made of modifications of conventional treatment systems to effect denitrification for nitrogen removal (43) and of physical-chemical treatment (chemical coagulation) for phosphorus removal (44–46). Such systems are unlikely to be applicable to individual household systems because of cost and complexity. At present, the only system that appears to be practical for nutrient removal is the shallow (aerobic) drainfield system, in which plant cover provides uptake of effluent nutrients as well as evapotranspiration (3, 32).

The uptake and transport of nutrients in soil is discussed further in the chapter, Treatment by Application onto Land.

4.4. Innovative Design

A Water for People project (47) resulted in a dual dry latrine system for flood-prone areas in Bolivia where frequent flooding washed out the contents of typical below ground level pit toilets. Dual chambers are built on a platform above the normal flood level and the chambers waterproofed. A special toilet seat is designed to separate urine from the feces. The urine is frequently used as a fertilizer. Ash or lime is applied to the solids in the chamber. After about 6 to 8 months use by a family of 6, the chamber is full, and the toilet seat is moved to the other chamber. By the time the second chamber is filled, the contents of the first chamber have decomposed to an ashlike material that can be applied to nonedible plants or trees. This system completely eliminates the need for treatment and disposal of the liquid typical of a water-carried waste.

4.5. Maintenance

Whereas septic systems are designed for minimum maintenance, they still require some maintenance. The septic tank itself is designed for periodic cleaning, normally by a professional septic tank cleaning service. All the tanks are equipped with a cleanout cover on the top. The homeowner should be familiar with its location. Although the original placement of the tank is normally located for convenient access, subsequent construction may constrict this access. A rule of thumb for locating unknown tanks in temperate climates is that the snow always melts fastest over the septic tank. This is due to the heat of the of the water drained into the tank and the bacterial activity within the tank.

Periodic cleaning of the septic tank is the secret to the longevity of the entire system. Relative to the leach field, the tank is the less expensive part of the system. If the tank is

cleaned before solids are carried over to the leach field, much more costly rebuilding of the leach field can be avoided.

The frequency of the need for cleaning of the septic tank depends upon not only the number of people using it, but also expeditious control of what goes down the drain. Obviously such non-decomposable materials such as glass, metals, and plastic should not be allowed to reach the septic tank. Ground food wastes are acceptable, usually with a larger capacity, but bones should not be included. Cloth, particularly diapers, should not go down the drain. Feminine sanitary items and wet strength toilet paper and towels should be excluded. Cigarette butts should not be thrown in the toilet. Overuse of anti-bacterial substances should be avoided, as these may destroy the balance of bacteria responsible for the breakdown of the organic matter in the tank. Similarly strong cleaning agents and drain cleaners could interfere with this bacterial activity. A reasonable amount of salt from water softening units can be tolerated, but this could be disposed of separately as with gray water. There is increasing concern that personal health care products and pharmaceuticals should not be allowed to enter the waste system because they may not be broken down in the treatment system and subsequently appear in a water supply at some later point. Once the household determines the necessary frequency of cleaning, it should establish a set time for repeat cleanings.

4.6. Restoration

Even with the most judicious maintenance of the septic tank, the leach field may not last forever. Some solids may be carried over and clog the soil. Biological growth will form mats that also cause clogging. Replacement of an entire leach field is expensive, so restoration alternatives are welcome. Some work and others are a waste of time and money.

US EPA has issued a fact sheet for Renovation/Restoration of Subsurface Wastewater Infiltration Systems (SWIS) (48). This reviews the various methods and products that have been developed for restoring the infiltration capacity of an SWIS.

Probably the most effective remediation is resting. Just as the daily resting period that occurs with normal household dosing, longer resting allows bacteria to mineralize the organic matter, thus unclogging the soil. Longer resting of 6 to 12 months has been found to be effective. However, this requires an alternative means of disposal of the liquid. It is suggested that a portion of the seepage area be rested, such as with a distribution box. However, if the entire SWIS is clogged, this will present a problem.

Studies have been explored of the potential for adding earthworms to open up the soil, but no conclusive data have been secured as of this time.

Many additives are available on the market claiming to improve the operation of a septic tank and/or restore a clogged SWIS. Strong acids or bases or toxic chemicals are generally discouraged. They may improve conditions in the septic tank, but pass on through and harm the infiltrative soil. Enzymes may increase liquefaction, but have not been shown to improve a clogged SWIS. Bacterial additives may speed up the start of the biological degradation in the septic tank, but all the needed bacteria are already available in animal fecal matter.

Hydrogen peroxide has been used to speed up the biological activity in a clogged SWIS and open the soil. However, long-term studies have shown that this is only temporary and it may permanently reduce the soil porosity and hydraulic conductivity.

An unproven system has been designed involving soil fracturing. It involves placing a steel tube into the soil below the infiltration surface and lifting the soil with high-pressure air. Polystyrene beads are then placed into the voids, holding them open as the air is released. The cost of this has not been evaluated.

5. CONCLUSIONS

Selection of the type of individual household wastewater disposal system depends on many factors, predominately site size and suitability for the proposed system (48). Attention should be given to the problem of potential nutrient pollution on sites close to surface water bodies. Assuming more than one system is acceptable, economics should determine the optimal choice (2, 17). Where suitable, well-designed septic tanks and tile fields provide the best system in terms of cost, proven reliability and low maintenance. On sites of limited size or poor soil permeability, aerobic systems may be favored because of their lower area requirements. Future development is likely to make aerobic systems more competitive in cost and more reliable. On particularly difficult sites, an innovative system utilizing sand filtration or evapotranspiration for effluent disposal may be necessary.

6. COST ESTIMATION

Precise calculation of costs for individual treatment systems is impossible, because such costs are dependent to a great extent on local conditions, including availability of equipment, construction costs, and land value. Another factor to be considered is the availability of federal and state construction grants, and state revolving funds (SRF) for financing at low interest rates. However, a general comparison of costs of anaerobic and aerobic systems can be made based on typical values (2, 6, 49)

A major cost of an anaerobic tank is the installation. Larger septic tanks cost relatively little more than a just acceptable sized tank. The difference in cost is readily made up in less frequent pumping out, and better protection of the drainage field. Further, if a family grows, the available additional volume will avoid the necessity of replacement of the tank with a larger size. The expected lifetime of a septic tank is 25 years. Thus the added cost of the next size larger tank will have only a negligible increase in the annual cost of the system.

Similarly, at a slightly additional cost, a two-compartment tank can significantly reduce clogging of the drain field. Having the second compartment available for inspection will signal the time for pumping before solids are carried over into the drain field.

Except for occasional pumping out and routine inspection for integrity, there is no operational cost for a septic tank.

In contrast, aerobic tanks have a much higher initial cost, have a constant cost for power (usually electricity), and still require occasional pumping out to maintain an optimal sludge balance. As a general rule, an aerobic tank installed will cost about twice as much as an anaerobic tank for a similar location. Further, they have an estimated lifetime of only about 10 years. Frequent inspection and maintenance are essential.

On the other hand, the drainage field area for an aerobic tank is about half that of an anaerobic tank. The value of the land required for the drain field represents a major cost. Thus it may

Table 3.7
Hypothetical US EPA rural community technology costs[a] (50)

Technology option	Total capital cost	Annual O&M cost	Total annual cost	Average monthly cost/household
Centralized system	$2,585,600 to $4,176,590	$33,110 to $44,830	$241,480 to $381,410	$149 to $235
Septic tank, small gravity sewer, cluster treatment	$666,040	$8,120	$61,800	$38
Onsite system	$567,940	$14,920	$60,690	$37

[a] All costs are in year 2000 US Dollars. Use Appendix A to calculate prices in year 2005 US Dollars.

be seen that the cost of the land can have a significant impact on the overall cost of a system. Further, in some instances not enough land may be available for an anaerobic tile field. In such a situation, cost may not be the limiting factor. A properly designed tile field should last for 25 years if precautions are made to prevent sludge from reaching it. Because replacement can be expensive, it is recommended that proper design is adhered to in the initial installation.

It is apparent that determining the cost of a system is very site specific. Many factors, including operational costs, maintenance, and the impact of power failure, must be considered in the final choice.

The US EPA has conducted a cost comparison (50) of conventional centralized collection and treatment, septic tanks followed by small diameter gravity sewers and cluster treatment, and onsite septic tank and leach field systems for a small community of 450 people in 135 homes. The results are summarized in Table 3.7. The study was conducted in 1997, and the table shows costs extrapolated to 2000. It may be seen that there was little difference in the costs between the two systems utilizing septic tanks, but these were considerably less expensive than the conventional centralized system.

It should be noted that estimates for design flow vary considerably from about 400 to 800 L/bedroom-d (100 to 200 gal/bedroom-d). In designing a treatment system, low capital cost differential and lower maintenance favor the most conservative design. In the case of the tile field, however, because peak loads are not likely to continue over extended time periods, the conservative design approach, though essentially guaranteeing trouble-free operation may result in consuming considerable excess area and money. Thus, in Table 3.5 criteria are included for the most conservative as well as moderately conservative (based on lower flow) designs. Judgment based on site suitability (e.g., soil permeability) and cost differential should be exercised. The readers are referred to the literature (51–64) for additional technical information on the subject.

7. SAMPLE DESIGN PROBLEMS

A three-bedroom home requires an individual wastewater disposal system. Assume level ground, with a percolation rate in a 15 cm hole of 10 min/cm (about 25 min/in.). Assume

sufficient suitable area for drainfield and required setbacks. Seasonal high ground water table is 2.3 m (7.5 ft) below the surface.

1. ***Determine design flow***: Estimate flow at 400 Lpcd (about 100 gpcd). Estimate occupancy at two persons per bedroom in the first two bedrooms, one and one-half in the additional bedroom, or 5.5 persons. Design flow = 2200 L/d (550 gpd).
2. ***Assume an anaerobic treatment system is selected***.
   a. ***Determine the required septic tank capacity***: The minimum tank capacity would provide 1 d detention time plus about 40% sludge and scum storage, or a capacity of about 3100 L (770 gal). However, for reasons previously discussed, the minimum design is not the most economical in the long run. A tank of at least 4800 L (1200 gal) capacity should be used (see Table 3.3). This represents an increase of 35% in the detention time and sludge and scum storage. The tank should be concrete, with two or three compartments.
   b. ***Determine the major drainfield parameters***.
   The percolation test indicates that the soil is suitable for a drainfield, and the safe loading rate of 11 L/d/m^2 should be used. The groundwater table indicates that the total trench depth should be no greater than 1.3 m (4 ft) to insure 1 m (3 ft) clearance above the water table. To maximize the effective sidewall area per unit length of trench, 1 .3 m (4 ft) trenches with 0.3 m (1 ft) backfill will be used. This provides 2 m^2 of effective sidewall area per meter of trench length (6 ft^2/ft). Based on conservative design (flow of 2200 L/d), the total trench length required is:

$$(2200\,\text{L/d})(1\,\text{m length})/(11\,\text{L/d/m}^2)(2\,\text{m}^2\text{area}) = 100\,\text{m length}$$

   The total trench length may be divided, for example, into four sections of 25 m (82 ft) length each, or six sections of 16.7 m (55 ft) length. Series distribution is used, and successive sections are separated by undisturbed ground equal to twice the depth (2.6 m or 8 ft). Trench width is 0.3 m (12 in.). One possible layout satisfying these requirements would require a total area required of about 36 m by 5.5 m (about 115 ft by 18 ft) or 198 m^2 (2100 ft^2). A similar area should be reserved for a replacement drainfield, should one become necessary.

   Note that the data presented in Table 3.5 produce a similar design. For three bedrooms, sidewall area required is:

$$(2\text{ rooms})(75\,\text{m}^2/\text{room}) + (1\text{ room})(55\,\text{m}^2/\text{room}) = 205\,\text{m}^2\text{ sidewall}$$

$$(205\,\text{m}^2)(1\,\text{m length})/(2\,\text{m}^2\text{ sidewall}) = 102.5\,\text{m total trench length}$$

   For the moderately conservative approach, (3 bedrooms) (55 m^2/room) (1 m length)/ (2 m^2 sidewall) = 82.5 m total trench length may be used in layout determination. Further details of trench design are outlined in the design section.
3. ***Assume an aerobic treatment system is selected.***
   a. ***Determine the required capacities of the tank chambers and the minimum air supply***.
   Design flow is 2200 L/d (500 gpd) as above.
      i. Provide an aeration chamber of 18-h detention time:

$$(2200\,\text{L/d})(0.75\,\text{d}) = 1650\,\text{L}\,(436\,\text{gal})$$

      ii. Provide a settling chamber of 6-h detention time:

$$(2200\,\text{L/d})(0.25\,\text{d}) = 550\,\text{L}\,(145\,\text{gal})$$

iii. Provide a fail-safe filter of 1.5-h detention time

$$(2200\,\text{L/d})(1.5\,\text{h})(1\,\text{d})/24\,\text{h} = 137\,\text{L}\,(36\,\text{gal})$$

iv. Minimum air supply:
Design based on 5.5 person occupancy, BOD removed:

$$(5.5\ \text{persons})(0.12\,\text{kg/cap-d}) = 0.66\,\text{kg BOD/d}$$

Air required:

$$(0.66\,\text{kg BOD/d})(150\,\text{m}^3\ \text{air/kg BOD}) = 100\,\text{m}^3/\text{d} = (3{,}530\,\text{ft}^3/\text{d})$$

This design is approximate, and should serve as a guide to selection of a unit. Details of construction will vary with the manufacturer.

b. Determine the major design parameters for a suitable aerobic seepage bed.
From Figure 3.16 for soil with a percolation rate of 10 min/cm (25 min/in.) a seepage rate of $22\,\text{L/d/m}^2$ ($0.54\,\text{gpd/ft}^2$) may be used. (This represents evapotranspiration in northern United States; upward adjustments may be made in some areas (3)).

i. Area required for seepage bed:

$$2200\,\text{L/d}/(22\,\text{L/d/m}^2) = 100\,\text{m}^2 = (1080\,\text{ft}^2).$$

Note that if the bed is designed to operate on infiltration alone (no evapotranspiration), the seepage rate (Figure 3.16) would be $18\,\text{L/d/m}^2$, and the required area $122\,\text{m}^2$ ($1320\,\text{ft}^2$).

ii. Any layout satisfying the areal requirement is suitable. The area required is modest for a single seepage bed. Such a bed may have dimensions of 6 m by 17 m (20×55 ft). To satisfy water and air distribution, four length-wise parallel 10 cm (4 in.) perforated pipes are suggested. Because the bed is charged with aerobic effluent, extra aeration pipes are optional. Further details of seepage bed construction are outlined in the design section.

4. Compare the corresponding anaerobic and aerobic units.
The required septic tank has a capacity of 4800 L (1270 gal), whereas the aerobic unit totals about 2400 L (634 gal). However, the aerobic unit requires an air supply and sludge recycle mechanism, and may be less stable in operation and require more maintenance.

The aerobic seepage bed requires about one-half the area of the anaerobic tile field ($100\,\text{m}^2$ vs. about $200\,\text{m}^2$) in this case. In addition the aerobic field may be more resistant to failure due to clogging, and exhibits better nutrient removal. Difficulties in providing a level infiltrative surface and cost of sand fill may somewhat offset savings in area requirements.

5. Design a suitable seepage pit for this three-bedroom house.
Seasonal high ground water table is 5 m (16.8 ft) below the surface. Two soil strata are present in the top 5 m (16.5 ft). From the surface to a depth of 2 m (6.6 ft), the soil has a percolation rate of 10 min/cm (25 min/m.), and from a depth of 2 to 5 m (6.6 to 16.5 ft), a rate of 6 min/cm (15 min/in.).

a. Determine the weighted average for the percolation rate:
The depth of the pit to be used must first be determined. Because a clearance of 1.3 m (4 ft) is required from the bottom of the pit to the high ground water table, the maximum allowable depth is 3.7 m (12 ft). Because this is a reasonable depth, design will be based on it. Thus, the weighted average percolation rate, R is:

$$R = \frac{[(10\,\text{min/cm})(2\text{m}) + (6\,\text{min/cm})(1.7\,\text{m})]}{(3.7\,\text{m})}$$

$$R = 8.16\,\text{min/cm (about 21 min/in.)}$$

b. ***Determine the dimensions for the percolation rate*:**
From Table 3.6, for a percolation rate of 8 min/cm, a sidewall area of about 19.5 m^2 per bedroom is required. Thus the total sidewall area required is about 59 m^2 (642 ft^2), where A is sidewall area, D is diameter, and d depth

$$A = \pi Dd \text{ or } D = A/\pi d \tag{4}$$

In this case

$$D = 59\,m^2/(3.14)(3.7\,m) = 5.08\,m\,(16.7\,ft)$$

Because this is an unreasonably large diameter, two pits in parallel are recommended, each with a diameter of 2.54 m (8 ft) and a depth of 3.7 m (12 ft), and thus each with a sidewall area of 29.5 m^2 (321 ft), i.e., one-half the total required area. The two pits must be separated by a distance equal to three times the diameter, or 7.6 m (25 ft).

NOMENCLATURE

A = cross-sectional area, m^2 (ft^2)
ASCE = American Society of Civil Engineers
BOD = biochemical oxygen demand, mg/L
D = diameter, m (ft)
DO = dissolved oxygen
EPA = Environmental Protection Agency
ET = Evaportranspiration
FHA = Federal Housing Administration
H = hydraulic head, m (ft)
K = hydraulic conductivity
L = cell height, m (ft)
P = tank effluent pollution load, mg/L
Q = volumetric flow rate through the sample cell, m^3/d (gpd)
Q = wastewater flow rate to seepage bed, L/d (gpd)
R = weighted average percolation rate, min/cm
S = hydraulic slope
s = seepage rate, L/d/m^2 (gpd/ft^2)
SS = suspended solids, mg/L
SWIS = Subsurface Wastewater Infiltration Systems

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APPENDIX

United States Yearly Average Cost Index for Utilities– U.S. Army Corps of Engineers (63)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

4
Submerged Aeration

Jerry R. Taricska, Jerry Y. C. Huang, J. Paul Chen, Yung-Tse Hung and Shuai-Wen Zou

CONTENTS

Abstract Submerged aeration is created by the action of submerged aerators (air diffuser, static mixer, sparge turbine, jet aerator, etc.), which introduce compressed air or oxygen at or near the bottom of an aeration basin. Oxygen transfer and liquid mixing are achieved as air/oxygen bubbles rise to the water surface. This chapter discusses the topics of aeration performance, oxygenation, deoxygenation, oxygen saturation, submerged aeration systems, aerators, oxygen transfer, design and applications.

Key Words Submerged aeration • performance • oxygenation • deoxygenation • aerators • air diffuser • static mixer • sparge turbine • jet aerator • oxygen transfer • design • application.

1. INTRODUCTION

Aeration is a mass transfer process by which oxygen molecules are exchanged between water and oxygen molecules at a gas/liquid interface. Aeration plays an important role in the purification of wastewater; the aeration process transfers oxygen to the wastewater and mixes the liquid contents such that the prevailing environment in the aeration basin permits microorganisms to use the organic material as a substrate for growth and a source of energy.

Depending on the location at which air or oxygen is introduced into the bulk of the water, aeration can be classified as either surface aeration or submerged aeration. In the former,

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oxygen transfer into the liquid phase is achieved by the contact of air and mist resulting from a hydraulic pump and/or the spray created by the rotation of an aerator near the surface of the water. (See "Surface and Spray Aeration" chapter.) In the latter, air or oxygen is released from orifices at a submerged depth and oxygen is transferred into the water during the course of air bubbles being generated and rising to the surface. This chapter deals with submerged aeration.

Submerged aeration is created by the action of submerged aerators such as air diffusers, static mixer, sparge turbine, and jet aerators. The aerators introduce compressed air or oxygen at or near the bottom of basin and have an oxygen transfer efficiency of 2.5 to 4.5 lb/hp-h (1.53 to 2.75 kg/kw-h) at standard conditions. Ability to maintain high liquid temperature and deep-tank application are the advantages. However, high initial cost and, in some cases, high maintenance cost is the disadvantages of these aerators.

Because the theory of gas transfer and aeration has been treated elsewhere in this book, it will not be considered here. (See the chapter on "Surface and Spray Aeration.") Instead, the more practical area of performance evaluation will be dealt with, performance characteristics of typical submerged aerators will be described, and a typical case study dealing with submerged aeration will be used as an illustrative design example.

2. AERATION PERFORMANCE EVALUATION

Three major factors to be considered in evaluating an aeration system are the efficiency in terms of oxygen transfer, the extent of mixing, and the flexibility in operation. (Also see the chapter on "Surface and Spray Aeration.")

There are two terms commonly used interchangeably in describing the efficiency of a submerged aeration system: oxygen absorption efficiency and oxygen transfer efficiency. These terms are not truly interchangeable, as can be seen from the following definitions. The former is defined as the percentage of oxygen transferred into the water out of the total amount of oxygen introduced into the water. No energy term is included. This term does not apply to surface aerators because no specific amount of air supplied through an air source such as an air blower can be determined in this type of aeration. The latter is defined as the rate of oxygen transferred into the water per unit of power consumed. The power is commonly based on the brake power unit, such as brake horsepower. Because oxygen transfer efficiency is applicable for both surface and submerged aeration systems, this term is commonly used as a basis for evaluating the efficiency of both types of aerators.

The efficiency of oxygen transfer is reflected in both initial and operating costs. The more efficient the aerator is, the lower total power requirement for the aerator, and, generally, the lower the initial equipment cost and operating cost.

Mixing is required to dampen the fluctuations of both hydraulic and organic loadings, to maintain uniform dispersion of the microorganisms, and to maintain uniform dissolved oxygen concentration in the mixed liquor contained in the aeration basin. Failure to provide proper mixing can result in the failure of the biological system.

Flexibility in operation is important in handling very wide fluctuations in hydraulic and waste loading conditions. The aerator is usually designed to provide oxygen to meet the demand of either projected organic loadings sometime in the future or expected peak oxygen

demands. For much of the time, the plant is usually operated at somewhat less than peak design conditions and does not require the total amount of oxygen that can be supplied. Therefore, an aeration system capable of handling wide ranges of oxygen demand can lead to substantial savings in operating costs and can be a major consideration in selecting the type of aerator to be used.

Although the basic aeration theory is generally universally accepted, there exist areas of disagreement not only in the details of conducting oxygen transfer tests, but also in the methods of data interpretation for evaluating aerator performance. Although several standards are in existence, these standards are concerned primarily with the methodology of experimental measurement and do not deal adequately with the interpretation and application of data to engineering design. Moreover, there is no general agreement among engineers and manufacturers about which standard to use. As a result, a wide variety of techniques is employed, resulting in substantial variations in results for the same device in the same testing conditions. Even larger variations are evident in translating these results to full-scale design. A vigorous effort has been made to develop a standard for evaluating oxygen transfer efficiency (1).

In evaluating aeration systems, four questions which confront engineers are:

1. Hydraulic regime (steady state or nonsteady state?).
2. Means of deoxygenation (use of chemicals or nitrogen gas?).
3. Oxygen saturation value (surface or mid-depth C_{ST}?).
4. Data interpretation.

2.1. Hydraulic Regimes of Performance Evaluation

There are two hydraulic regimes in evaluating aeration systems: steady state and nonsteady state.

2.1.1. Steady State

In aerating wastewater under steady state, the microbial oxygen uptake rate, gamma (γ), is measured in an operating activated sludge system in the absence of immediate oxygen demand and when there is more than 1 mg/L of dissolved oxygen level. The waste treatment system must be at constant steady-state conditions throughout the test period with the dissolved oxygen concentration, the organic loading, wastewater temperature, and mixed liquor suspended solids level all held constant.

It is difficult to control the many variables at a constant steady state; accurate determination of the microbial oxygen uptake rate, γ, is very difficult; and test results can not be compared unless they are corrected to standard conditions with regard to alpha (α), beta (β), and liquid temperature. (See the chapter on "Surface and Spray Aeration" for definitions of α and β factors.) This makes the test unacceptable as a universal standard evaluation method. It is, however, a very useful indicator of the relative performance of different aeration devices tested with the same waste in parallel basins.

In aerating clean water under steady state, deoxygenated water is continuously fed to an aeration basin at a constant flow rate. Complete mixing of the tank contents at constant temperature is required. This means that the system must be in a hydraulic steady state

throughout the testing period, which can be accomplished only by operating the aerator for a sufficient length of time. Additionally, high flow rates are required to obtain accurate results. This procedure limits the test to the evaluation of only small aerators in a relatively small aeration basin. The difficulty in maintaining a perfect steady state and its limitation to small aerators preclude the wide acceptance of this test as a universal standard evaluation method.

2.1.2. Nonsteady State

In the nonsteady state hydraulic regime, there are also two different cases.

- In aerating wastewater under nonsteady state, the test is used to determine the microbial oxygen uptake rate, γ, in activated sludge. Complete mixing of the tank is required during the aeration portion of the test. The aeration proceeds until an equilibrium DO (dissolved oxygen) level is achieved (preferably greater than 3 mg/L), after which aeration is stopped and the oxygen content of the liquid in the aeration tank is allowed to decrease as a result of microbial respiration until only a minimal amount of dissolved oxygen remains. There should be no significant change in microbial concentration or process loading during the test period. The oxygen transfer rate determined is related to the particular wastewater tested and, thus, restricts the test in terms of comparison to other tests. Additionally, it is difficult to keep the tank or basin contents homogeneous after turning off the aerators.
- A nonsteady state clean water evaluation involves the reaeration of clean water that has previously been depleted of dissolved oxygen by either sodium sulfite or nitrogen gas. A completely mixed tank and a uniform water temperature are required during the test. This test is controlled by interfacial diffusion and is not complicated by biological reactions. This nonsteady state clean water hydraulic regime has been adopted by the PEMA (Process Equipment Manufacturer's Association) (2) as a standard hydraulic testing condition for the comparison of oxygen transfer rate and efficiency by different types of aeration equipment.

Eckenfelder (3), Morgan and Bewtra (4), and Bewtra and Nicholas (5) have reported on the importance of the test basin geometry. It is generally recognized that basin geometry, aerator placement, and power level affect oxygen transfer performance. Normally, one attempts to closely simulate field installation conditions for specific applications.

PEMA (2) recommended these criteria for diffused aeration:

Basin size:	A rectangular test tank: minimum length = 1.5 × tank width and depth = 3–6 ft (0.914–1.83 m)
Air supply:	12 to 45 scfm/1000 ft^3 (0.012 to 0.045 m^3/min -m^3)
Diffuser submergence:	8 to 14 ft (2.44 to 4.27 m)
Power level:	0.05 to 0.20 hp/1000 gal (0.00985 to 0.0394 kw/m^3)

2.2. Means of Deoxygenation

Nitrogen stripping and sodium sulfite are the two methods commonly employed. Nitrogen stripping is primarily used in laboratory and some shop testing (6, 7). This method exhibits the merits of not increasing dissolved solids in the testing water and of avoiding the chemical interference in oxygen determination by the Winkler method.

The theoretical requirement of sodium sulfite ($NaSO_4$) is 7.88 mg/L of sulfite per 1.0 mg/L DO concentration. Sulfite additions are made in excess and are dependent on the oxygen transfer rate and mixing rates. Standard methods (8) and Manual of Practice (9) suggest 0.8 lb of sodium sulfite/1000 gal(0.096 kg/m^3), which is 22% excess. PEMA (2) suggests 1 lb of sodium sulfite/1000 gal(0.120 kg/m^3), which is 50% excess.

The rate of sulfite oxidation with DO is very slow and can take hours before all of the sulfite ions are oxidized (4, 10). If the unreacted sulfite ions are still present during the subsequent oxygenation step, an erroneously low aerator oxygen transfer capacity will result. To accelerate the kinetics of the sulfite oxidation process, addition of cobaltous ions as a catalyst has been practiced. PEMA (2) recommends a cobalt chloride or sulfate of between 0.1 and 2.0 mg/L. A higher 1evel of cobaltous ions to a concentration of 2 to 5 mg/L, and even as high as 10 mg/L, has become accepted in the practice of aerator eva1uations (11).

However, it been reported that using more than 0.05 mg/L cobaltous ion concentration in the catalyzed deoxygenation process with sodium sulfite has caused interference in the Winkler method of DO determinations (11–13). The degree and magnitude of this interference vary with the cobalt ion concentration, the amount of sulfite additions, pH, and the buffering capacity of the test water. Naimie and Bums (13) reported that the interferences are caused by mixed hydrated trivalent and tetravalent cobalt oxides and hydrogen peroxide. They postulated that the interference would probably be eliminated if the aeration test were conducted at pH values of 6.9 or below, and suggested an addition of 0.001 M phosphate at buffer pH 6.9 before actual testing. The cobalt interference in a soft water system is minimal or is totally eliminated; however, water systems with high pH and alkalinity will cause various unpredictable magnitudes of interference at the practiced cobaltous ion concentration range of 2 to 5 mg/L.

It has been common practice to limit the number of runs on a given volume of water, under a certain upper limit concentration of either TDS (total dissolved solids) or sulfate. Landberg et al. (12) noted an effect on the Winkler DO determination at 1.0 mg/L cobalt concentration and a Na_2SO_4 concentration of about 1200 mg/L. Other investigators (2, 14, 15) reported an upper limit of sulfate concentrations ranging from 1000 to 1350 mg/L. These references do not make any distinction on upper limits if probe analysis for DO concentration is used. Some testing has been conducted at higher TDS concentrations without apparent reproducibility problems.

2.3. Oxygen Saturation Concentration

In submerged aeration, oxygen mass transfer occurs throughout the volume and oxygen saturation concentration, C_{ST} varies with depth because of progressive decreases in both hydrostatic pressure and oxygen mole fraction as the gas phase moves upward. Because the accuracy with which the oxygen transfer rate can be predicted depends greatly on the accuracy with which C_{ST} can be determined, the pivotal factor in evaluating the performance is the value of C_{ST} used to determine the oxygen deficit, or driving force.

Several studies (16, 17) have modeled submerged aeration based on the assumptions that:

1. The overall mass transfer coefficient, K_La, is constant over the tank volume.
2. Good mixing exists so that the DO concentration, C is uniform over the tank volume.
3. Oxygen is the only gas transferred.

In the case of uniformly distributed submerged bubble aeration, these parameters might not vary appreciably. However, where air input is nonuniform or where a circulatory motion or turnover is induced in the tank, $K_L a$ and C_{ST} could vary significantly over the tank liquid. It should be noted that a spatial variation in dissolved oxygen concentration could exist and it significantly influences the oxygen transfer.

In some instances C_{ST} has been assumed constant and equal to the surface saturation value. Stanton and Bradley (18) have shown that this assumption seriously underestimates the true saturation concentration and yields inflated values of $K_L a$.

The most satisfactory model assumes the value of C_{ST} to be the arithmetic average of the theoretical oxygen saturations at the point of bubble formation near the basin bottom and at the basin surface where the air bubbles escape to the atmosphere. It also takes into consideration the average concentration of oxygen in the air bubble during contact. In 1956, Oldshue (19) proposed all equation relating C_{ST} to oxygen content of the exit air, hydrostatic pressure, and the surface saturation value:

$$C_{ST} = C_S \left[\frac{F_i}{2F_o} + \frac{P_a + \gamma_w Z_d}{2P_a} \right] \tag{1}$$

where C_{ST} is the middepth oxygen saturation concentration at testing water temperature T, but corrected for water depth and partial pressure, C_S is the saturation concentration in pure water of temperature T under test site barometric pressure, F_i is the mole fraction of oxygen in the exit air, F_o is the mole fraction of oxygen in the feed air, P_a is one atmosphere pressure in psi, γ_w is weight density of water, Z_d is aerator submergence. According to Henry's law, in which $C_S = H F_o P_a$, where H is Henry's constant, Equation (1) becomes:

$$C_{ST} = \frac{H}{2} \left[F_o \left(P_a + \gamma_w Z_d \right) + F_i \, P_a \right] \tag{2}$$

Downing and Boon (16) in 1960 and Lister and Boon (20) in 1973 applied an equation slightly different from that of Oldshue to the analysis of submerged aeration. Their equation is in the form of:

$$C_{ST} = \left(\frac{F_i + F_o}{2} \right) \left(P_a + \frac{\gamma_w Z_d}{\varphi} \right) H \tag{3}$$

where φ is a constant characteristic of the aeration bubble pattern.

A third relationship was proposed by Baillod (21), assuming that the effective average saturation concentration corresponds to the arithmetic average of the feed air exposed to a pressure at 2/3 the depth and exit air exposed to a pressure at 1/3 of the depth. His equation is:

$$C_{ST} = \frac{H}{2} \left[F_o \left(P_a + \frac{2}{3} \gamma_w Z_d \right) + F_i \left(P_a + \frac{1}{3} \gamma_w Z_d \right) \right] \tag{4}$$

It is interesting to note that C_{ST} estimated values from Equations (3) and (4) are almost identical, but are slightly lower than that from Equation (2).

As early as 1934 investigators (22, 23) found that diffused aeration equipment furnished a dissolved oxygen saturation value equal to the saturation of approximately 1/3 the depth from

liquid surface to the diffusers. Other investigators (24–26) have noted a significant difference between measured and calculated C_{ST}.

2.4. Data Analysis and Interpretation

The proper choice of C_{ST} value plays an important role in determining an accurate overall oxygen transfer coefficient. $K_L a$, because they are related in the widely accepted two-film theory of mass transfer of oxygen from the gas phase to the liquid phase as follows:

$$\frac{dC}{dt} = K_L a\,(C_{ST} - C) \tag{5}$$

In most cases, $K_L a$ and C_{ST} are assumed to be constant over time. By specifying initially that $C = C_o$ at $t = 0$, Equation (5) can be integrated to yield:

$$\ln(C_{ST} - C) = \ln(C_{ST} - C_o) - K_L a \cdot t \tag{6}$$

or

$$C = C_{ST} - (C_{ST} - C_o)\exp(-K_L a \cdot t) \tag{7}$$

Most nonsteady state aeration tests are performed in such a manner that the bulk oxygen concentration, C, is measured as a function of time. Generally C is assumed to be uniform over the tank volume. These data are then modified to the form of Equations (5–7) so that estimates of the parameters C_o, $K_L a$, and/or C_{ST} can be obtained. Though C and t are the data to which the equations are fitted, other variables are measured and/or controlled during a test. These variables include temperature, relative humidity, and barometric pressure, which allow correction to standard conditions. Other variables such as air flow rate, exit air composition, tank geometry, and so on are also recorded to determine their effect on transfer.

It has been shown that an error in the value of C_{ST} will cause data to plot in other than a straight line on the semilog plot of oxygen deficit vs. time (24, 25, 27). If the value of C_{ST} used in the calculation of dissolved oxygen deficit is higher than the true value of C_{ST}, the plotted line tends to curve upward as the oxygen deficit decreases in the plot of deficit vs. time. When the calculated value of C_{ST} is lower than the actual value of C_{ST}, the plotted line tends to curve downward as the oxygen deficit decreases (24, 27).

To avoid the problem of nonlinear plots, many investigators (12, 15, 24, 28) recommend truncation of the data at 70% to 90% of C_{ST}. However, as shown by Boyle et al. (29), this is a questionable practice. Truncation can introduce considerable error in estimating $K_L a$. The data will appear linear, but the slope will be sensitive to the assumed value of C_{ST}. In other words, truncation increases the correlation between $K_L a$ and C_{ST}, but it reduces the precision of the estimated $K_L a$. Therefore care must be exerted in truncating data at the higher end.

In many cases dissolved oxygen data at the beginning of a reaction test are less reliable than observations made at higher concentrations. It can be discarded below 10% to 20% of saturation without affecting the precision of estimating the overall oxygen transfer coefficient, $K_L a$ (29).

For the purpose of comparing aerators' oxygen transfer efficiency under a common basis, $K_L a$ must be corrected to standard temperature, 20°C. The appropriate correction has been

found empirically to be

$$K_L a_{20} = K_L a_T \, (\theta^{T-20}) \tag{8}$$

where T is the test liquid temperature in °C; and θ is a constant for which values of 1.020 to 1.028 are normally used in submerged aeration systems. The amount of oxygen transferred to the test water at standard conditions, N_o, i.e., 20°C, maximum oxygen deficit ($C = 0$), and 1 atm of pressure can be calculated as:

$$N_o = K_L a_{20} (C_{SS}) W \tag{9}$$

where C_{SS} is the middepth oxygen saturation concentration corrected to standard conditions; and W is the weight of water being aerated.

The value of C_{SS} is related to:

$$C_{SS} = 9.17 \frac{C_{ST}}{C_S} \tag{10}$$

where C_{ST} and C_S are previously defined.

The oxygen transfer efficiency is defined as the oxygen transfer rate per unit power consumption. It is presented in the equation below.

$$E = \frac{N_o}{\text{Power}} \tag{11}$$

2.4.1. Example 1.

Calculate the overall oxygen transfer coefficient, $K_L a$, at test conditions and at 20°C, the oxygen transfer rate in lb O_2/h, and the aerator's oxygen transfer efficiency in lb O_2/hp-h, based on the results of a nonsteady state clean water reaeration test. Test conditions were: water temperature = 22°C, barometric pressure = 758.7 mm of Hg, basin volume = 2.2×10^6 gallons, air diffuser submergence = 17 ft, air flow rate = 2000 scfm, and blower wire-to-water horsepower = 177.1 bhp.

Reaeration data:

Time, min	4	8	12	16	20	24	28	32	36	40	44	48	52
DO, mg/L	0.1	2.1	3.5	4.5	5.6	6.2	7.0	7.4	8.0	8.3	8.7	8.8	9.1

(a) Determining MidDepth Oxygen Saturation Concentration, C_{ST}.

Because the dissolved oxygen saturation value in distilled water of temperature 22°C under 1 atm of pressure is 8.8 mg/L, then C_s at a pressure 756.7 mm of Hg is calculated as follows:

$$C_S = \text{DO}_{760} \left(\frac{P}{760} \right) \text{ at elevation less than 3000 ft.} \tag{12}$$

$$C_S = \text{DO}_{760} \left(\frac{P - p}{760 - p} \right) \text{ at elevation greater than 3000 ft.} \tag{13}$$

where p is the pressure (mm) of saturated water vapor at the temperature of the water. Assuming elevation less 3000 ft then

$$C_S = 8.8\left(\frac{758.7}{760}\right) = 8.78\,\text{mg/L}$$

Calculate an average F_i based on data between 4 and 52 min, disregarding the effect of surface reaeration.

$$\text{O}_2 \text{ rise in the basin} = 9.1 - 0.1 = 9.0\,\text{mg/L}$$

$$= \frac{\left(9.0\frac{\text{mg}}{\text{L}}\right)\left(2.2\times10^6\,\text{gal}\right)\left(3.785\frac{\text{L}}{\text{gal}}\right)}{\left(1000\frac{\text{mg}}{\text{g}}\right)\left(32\frac{\text{g}}{\text{g-mol}}\right)}$$

$$= 2.34\times10^3\,\text{g-mol}$$

A mole of any gas (air) occupies a volume of approximately $1.26\,\text{ft}^3$ (22.4 liters) then:

$$\text{air flow} = 2000\,\text{scfm}\,(48\ \text{min})\left(\frac{\text{g}-\text{mol}}{1.26\,\text{ft}^3}\right) = 0.76\times10^5\,\text{g–mol}$$

O_2 input owing to aeration amounts to 21% of the total air flow, which

$$0.76\times10^5\,\text{g–mol}\times0.21 = 1.6\times10^4\,\text{g–mol}$$

Therefore, the oxygen absorption efficiency, which is the fraction of oxygen in air bubbles transferred to testing water, is determined as follows:

$$e = \frac{2.34\times10^3\,\text{g–mol}}{16.0\times10^3\,\text{g–mol}} = 0.146$$

The mole fraction of oxygen in the exit air flow is:

$$F_i = \frac{21(1-e)}{79+21(1-e)} = \frac{21(1-0.146)}{79+21(1-0.146)} = 0.185$$

Therefore, the middepth oxygen saturation concentration is estimated, according to Equation (1), as

$$C_{ST} = 8.78\left[\frac{0.185}{(2)(0.21)} + \frac{14.7+(17\,\text{ft})(0.433\,\text{psi/ft})}{(2)(14.7)}\right] = 8.78\,[0.44+0.75] = 10.45\,\text{mg/L}$$

(b) Determination of the Overall Oxygen Transfer Coefficient.

Applying Equation (5), a semi log plot of $C_{ST}-C$ vs. time should yield a straight line with a slope of K_La. Figure 4.1 shows the data plotted for this testing data where the effective driving force is $C_{ST}-C$.

Three saturation values are used to show the effect of saturation values on the parameter estimated. In curve 1, the middepth oxygen saturation concentration, $C_{ST} = 10.45\,\text{mg/L}$ is used for determining the oxygen deficit; whereas in curve 2 the measured oxygen saturation concentration, $C_{ST} = 9.5\,\text{mg/L}$, is used; and in curve 3, the surface saturation value, $C_{ST} = 8.78\,\text{mg/L}$, is used.

Because the rate of oxygen transfer is known to be directly proportional to the DO deficit, a plot of the true deficit vs. time on semi log paper yields a straight line. However, as

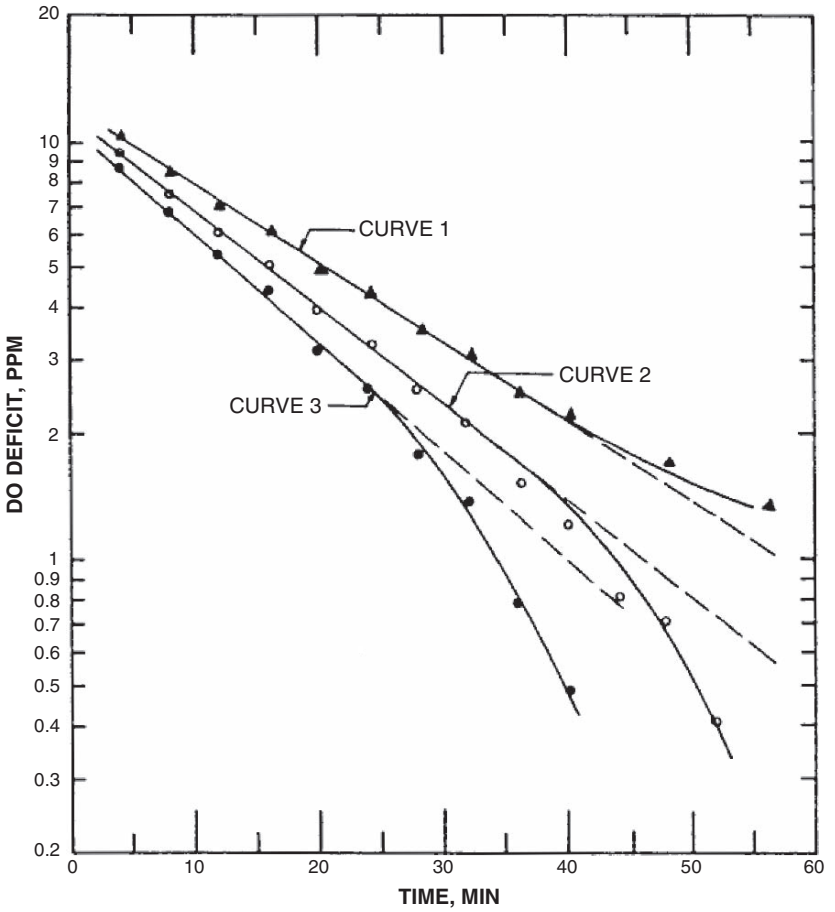

Fig. 4.1. Dissolved oxygen deficit vs time.

shown in Figure 4.1, the nonlinearity begins at 76%, 84%, and 75% of saturation values, for the curves 1, 2, and 3, respectively. An upward curve beyond 76% of saturation value indicates that the estimated middepth saturation value is slightly too high. On the other hand, the downward curves of curves 2 and 3 indicate that the observed and surface saturation values are too low. Therefore, it appears that the true saturation value lies between 9.5 and 10.45 mg/L.

The overall oxygen transfer coefficient, $K_L a$, is determined for curve 1 at $t = 5$ min and $t = 40$ minutes as:

$$K_L a = \frac{\ln 9.7 - \ln 2.1}{40 - 5} = 0.0437\,\text{min}^{-1} = 2.623\,\text{h}^{-1}$$

The temperature correction of the parameter is made according to Equation (8) as

$$(K_L a) = (K_L a)_{22} \times 1.024^{20-22} = (2.623)(1.024^{-2}) = 2.50\,\text{h}^{-1}$$

(c) Determination of Oxygen Transfer Rate and Efficiency: The oxygen transfer rate is estimated according to Equations (9) and (10)

$$N_o = (K_L a)_{20}(C_{SS})W$$
$$= (2.50\,h^{-1})\left(9.17 \times \frac{10.45}{8.78} \times 10^{-6}\frac{\text{lb } O_2}{\text{lb water}}\right) \times \left(2.2 \times 10^6\,\text{gal} \times 8.34\frac{\text{lb}}{\text{gal water}}\right)$$
$$= 501\,\text{lb } O_2/\text{hr}$$

Therefore, the aerator's oxygen transfer efficiency, E, is the result of N_o divided by the total wire-to-water horsepower used.

$$E = \frac{501\,\text{lb } O_2/\text{hr}}{177.1\,\text{bhp}} = 2.83\,\text{lb } O_2/\text{bhp} - \text{hr}$$

Applying the same procedure, a comparison can be made when different saturation values are used.

Parameters	Middepth saturation	Measured saturation	Surface saturation
Curve No. on Figure 4.1	1	2	3
C_{ST}, mg/L	10.45	9.50	8.78
$(K_L a)_{20}$, h^{-1}	2.50	3.08	3.44
N_o, lb O_2/h	501	619	692
E, lb O_2/hp-h	2.83	3.50	3.91

It is interesting to note that the oxygen transfer efficiencies for measured saturation and surface saturation are 23% and 37%, respectively, over that of middepth saturation. This illustrates the importance of specifying the conditions on which an evaluation and comparison of aerators are to be made.

3. SUBMERGED AERATION SYSTEMS

3.1. System Components

In a submerged aeration system, there are three major components: air supply source, air and/or liquid piping, and the aerator itself.

3.1.1. Blowers

The air is supplied by the air blower, which is one of two types: rotary positive displacement or centrifugal. The former type of blower is characterized by constant output of air flow with varying discharge pressure; the latter type of blower is characterized by relatively constant discharge pressure with varying output of air flow. Each type of blower has its unique application in submerged aeration. Aerating storm water that is temporarily stored in a reservoir or wastewater stored in equalization basin requires a type of blower that can produce a relatively constant air flow at various water pressure conditions. The rotary positive displacement blower is an excellent application in this case. This type of blower is also commonly used in deep aeration [water depth greater than 22 ft (9.5 psi)] application.

On the other hand, the centrifugal blower is used in cases where the oxygen demand of the waste stream varies considerably. When the oxygen demand is low the air supply is reduced by throttling the intake valve of the centrifugal blower. Thus, energy that would be wasted in producing excessive amounts of air is saved. Flexibility in operation is one of the important advantages that the centrifugal blower possesses. In general, the centrifugal blower is used in aeration that requires relatively low air pressure, but relatively large air flow rate.

In specifying blower performance, one must use actual cubic feet per minute (acfm). System requirements for oxygen are usually specified as standard cubic feet minute (scfm). Therefore, blowers must capacity must be determined at actual site conditions. Actual air requirement is determined as follows:

$$Q_{\text{acfm}} = Q_{\text{scfm}} \frac{P_S - (RH_S PV_S)}{P_B - (RH_A PV_A)} \left(\frac{T_A}{T_S}\right) \left(\frac{P_B}{P_A}\right) \tag{14}$$

where:

P_A = Actual pressure, psia
P_B = Atmospheric pressure, barometer psia
P_S = Standard pressure, 14.7 psia
RH_A = Actual relative humidity
RH_S = Standard relative humidity, 0.36
PV_A = Actual vapor pressure of water at actual temperature, psi
PV_S = Standard vapor pressure of water at actual temperature, 0.3391 psi
T_A = Actual temperature, °R (note °R = °F + 460)
T_S = Standard temperature, 528° R

3.1.1.1. Example of Blower Air Requirements

Assume the system requirements for diffuser system are as follows:

$Q_{\text{scfm}} = 500$ cfm
$P_B = 14.4$ psia
$P_A = 14.2$ psia (This assume 0.2 psia pressure drop ahead of the blower)
$T_A = 560°$ R(100° F)
$RH_A = 0.85$

Solution

Actual flow rate is determined from Equation (14).

$$Q_{\text{acfm}} = (500) \frac{14.7 - (0.36)\,(0.3391)}{14.4 - (0.85)\,(0.9503)} \left(\frac{560}{528}\right) \left(\frac{14.4}{14.2}\right)$$

$$Q_{\text{acfm}} = 576.8 \text{ cfm}$$

Therefore, the blower capacity would be sized to provide approximately 580 cfm at site conditions.

3.1.2. Piping

The requirement of air and/or liquid piping depends on the type of aerator. All submerged aerators require air piping from which air is transported to the submerged aerator from the onshore air blower. Additionally, liquid (wastewater) piping is required for the jet aerator in which the mixed liquor from the aeration basin is recirculated back through the jet aerator assembly. Other types of submerged aerators require no liquid piping component.

3.1.2.1. EXAMPLE AIR PIPING

Determine the head loss in 10 inch diameter air piping main with equivalent length of 150 feet in length and operating with a flow rate of 500 cfm (or 8.33 cfs).

Solution

$$\text{Velocity in } 10'' \text{ pipe} = \frac{8.33}{\pi \left(\frac{10}{12}\right)^2/4} = 15.27 \text{ fps}$$

Friction loss in 10″ pipe based on air 0.075 lb per cu ft density flowing through average, clean, round galvanized metal duct.

$$\frac{\text{Friction loss}}{100 \text{ ft}} = \frac{h_f}{100 \text{ ft}} = 2.74 \frac{\left[\frac{V_{fpm}}{1000}\right]^{1.9}}{[D_{inches}]^{1.22}} \tag{15}$$

$$\frac{h_f}{100 \text{ ft}} = 2.74 \frac{\left[\frac{(15.27)(60)}{1000}\right]^{1.9}}{[10]^{1.22}} = 0.140 \frac{\text{inches of water}}{100 \text{ ft}}$$

$$\frac{h_f}{100 \text{ ft}} = 0.140 \frac{\text{inches of water}}{100 \text{ ft}} \frac{\text{ft}}{12 \text{ inches}} = \frac{0.01167 \text{ ft}}{100 \text{ ft}}$$

Friction loss in 150 feet of 10″ pipe is

$$150 \frac{h_f}{100 \text{ ft}} = 150 \text{ ft} \frac{0.01167 \text{ ft}}{100 \text{ ft}} = 0.0175 \text{ ft} = 0.0175 \text{ ft} \left(\frac{0.433 \text{ psi}}{\text{ft}}\right) = 0.0076 \text{ psi}$$

3.1.3. Submerged Aerators

As far as the submerged aerators themselves are concerned, there is quite a variety of shapes and forms (30, 31). In general, submerged aerators can be classified into four major groups: diffused air aerator, sparge turbine aerator, static mixing aerator, and jet aerator as illustrated in Figure 4.2. Each group has its unique characteristics, advantages as well as disadvantages, and primary application area as summarized in Table 4.1.

3.2. Major Types of Submerged Aerators

3.2.1. Air Diffuser

Air diffusers bubble compressed air into water through orifices, nozzles in air piping, diffuser plates or tubes, or sparges. Diffused aeration equipment can be classified into two

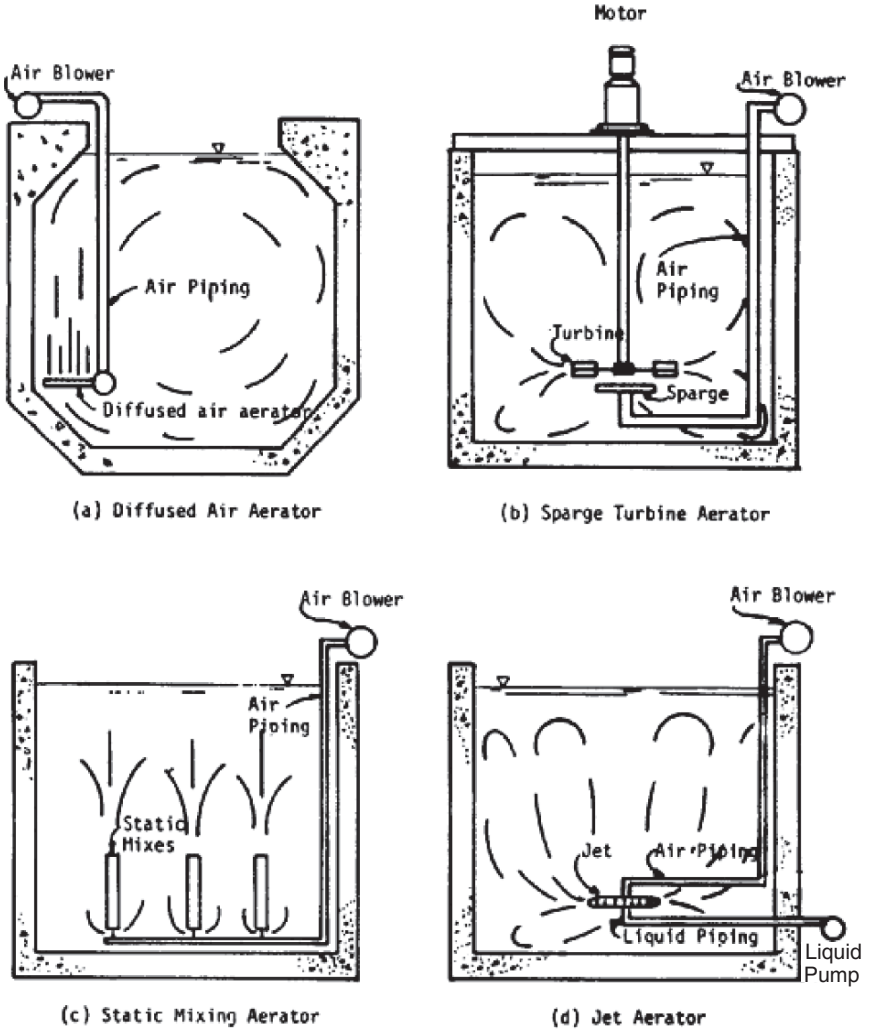

Fig. 4.2. Four major types of submerged aerators.

general types depending on bubble size generated. Large bubbles generated from coarse diffusers have the advantage of low maintenance over fine bubble diffusers. However, a lower oxygen absorption efficiency of 4% to 8% and oxygen transfer efficiency of 1 to 2 lb/hp-h (0.61 to 1.22 kg/kw-h) under standard conditions results from coarse air diffusers, whereas fine bubble diffusers generally have an oxygen absorption efficiency of 8% to 15% with an oxygen transfer efficiency of 2 to 3.5 lb/hp-h (1.22 to 2.13 kg/kw-h) under standard conditions.

The coarse air diffuser can be applied over a wide range of air flow rates as high as 25 cfm (0.708 m^3/min) per diffuser. Because of large orifice openings [1/8 to 3/8 in. (0.318 to 0.953 cm) diameter], higher air flow rates are attainable with small pressure loss. For example,

Table 4.1
Major aerators and their applications

Aerators	Characteristics	Advantages	Disadvantages	Primary applications
1a. Coarse air diffuser	Large air bubbles; from nozzle, valve, orifice, or shear plates	Nonclogging; low maintenance cost; no icing problem	Low oxygen transfer efficiency; high initial cost	All types of activated sludge processes
1b. Fine air diffuser	Fine air bubbles; from ceramic plates or tubes, plastic-wrapped or plastic-cloth tube or bag	High oxygen transfer efficiency; high mixing capability; no icing problem, no depth limitation	High initial and maintenance costs; tendency to clog	All types of activated sludge processes
2. Static mixing aerator	Short elements of right- or left-hand helices inducing inline mixing	Flexible in basin geometry; nonclogging	High initial and maintenance costs	Aerated lagoon and activated sludge processes
3. Sparge turbine aerator	Units contain a low-speed turbine and provide compressed air on sparge ring; fixed. bridge support	Deep-tank application; high mixing capability; no icing problems; upgrading of existing aeration facility	Moderate oxygen transfer efficiency; high maintenance requirement	All types of activated sludge processes and aerated lagoon
4. Jet aerator	Pump recirculates the mixed liquor and provides motive force; blower supplies compressed air to the jet assembly	High oxygen transfer efficiency and mixing capability; deep-tank application; upgrading existing aeration facility	High initial cost	All types of activated sludge processes and aerated lagoon

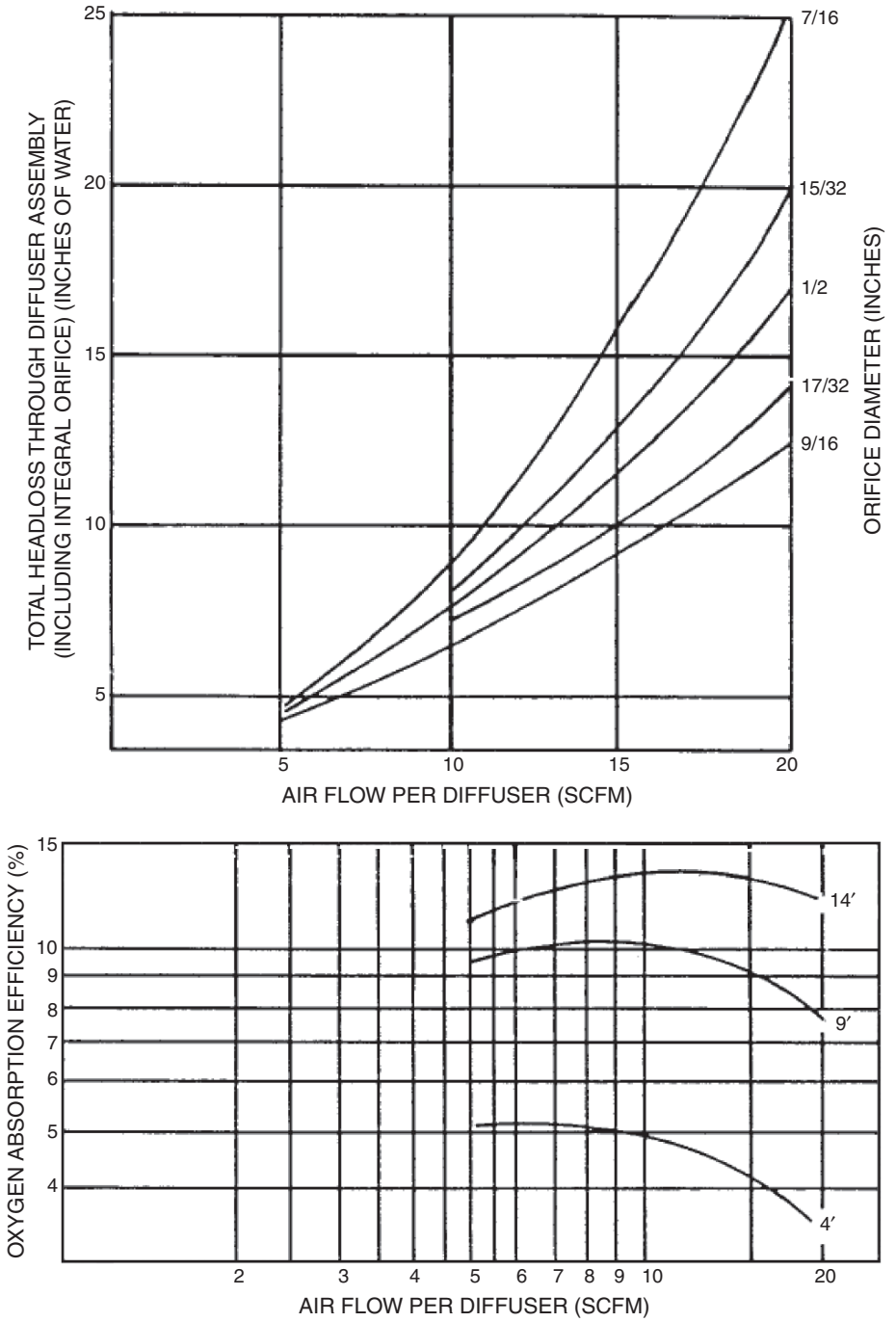

Fig. 4.3. Total headloss through diffuser and oxygen absorption efficiency as a function of the air flow rate.

a plastic diffuser may pass 10 to 15 cfm (0.283 to 0.425 m^3/ min) at a loss of pressure of 6 inches (15.24 cm) of water.

A standard fine air diffuser delivers from 5 to 20 cfm (0.142 to 0.566 m^3/min) of air per diffuser. To maintain adequate circulation in the aeration tank, aerators are placed so that they will deliver a minimum of 3 cfm/ft (0.279 m^3/min -ft) along the tank. Fine-bubble diffusers are subject to clogging either externally by solids in the liquid, or internally by particulate matter in the air supply. Therefore, routine periodic cleaning of the air diffusers or replacement of air diffusers is required.

Figure 4.3 shows the total head loss through diffuser and oxygen absorption efficiency as a function of the air flow per diffuser for one type of sock diffuser. The head loss increases exponentially as air flow per diffuser increases and this pattern is more profound as the orifice diameter of the diffuser decreases. There exists an optimum air flow per diffuser at which oxygen absorption efficiency is the highest for specific diffuser submergence. As the submergence increases, the air flow per diffuser that results in the optimum oxygen absorption efficiency shifts to a higher value. For example, the optimum oxygen absorption efficiency of 13.5% occurs when an air flow of 12 scfm (0.340 mL/min) per diffuser is provided for a 14 ft (4.27 m) submergence, whereas the optimum oxygen absorption efficiency of 5% occurs when an air flow of 6.5 scfm (0.184 ml/min) per diffuser is provided for a 4 ft (1.22 m) submergence. This figure also indicates that the oxygen absorption efficiency is very much diffuser submergence dependent—the deeper the diffuser is, the higher is the oxygen absorption efficiency.

However, the oxygen absorption efficiency is not always the best criterion in evaluating a diffuser. In some instances, a diffuser can reach high absorption efficiency at the expense of high power requirement. Therefore, the power requirement for a diffuser to achieve a certain absorption efficiency and oxygen transfer rate is another important criterion to be evaluated.

Figure 4.4 shows typical power requirements for the air diffuser at various air flow rates assuming an oxygen absorption efficiency of 6%. For a fixed amount of air flow rate, there will be a corresponding oxygen transfer rate (lb O_2/h) for assumed oxygen absorption efficiency. It should be noted that the power requirement is diffuser submergence-dependent: the deeper the submergence is, the higher the discharge air pressure required and the higher the power requirements.

3.2.2. Sparge Turbine Aerator

The sparge turbine aerator consists of a combination of a submerged turbine aerator having a rotating impeller, and a stationary diffuser ring for the injection of compressed air. Air rising from the diffuser ring is dispersed by the impeller and distributed throughout the liquid. Oxygen is transferred by air bubbling at the sparge ring, as well as by the shearing action on the bubbles by the impeller located above the ring. Upper impellers are often used to set up auxiliary mixing patterns.

Sparge turbine performance can be controlled by adjusting the ratio of turbine mixing energy to sparged air energy. These units generally have higher oxygen transfer efficiencies and energy use efficiencies than diffusers, but somewhat less than low speed surface aerators.

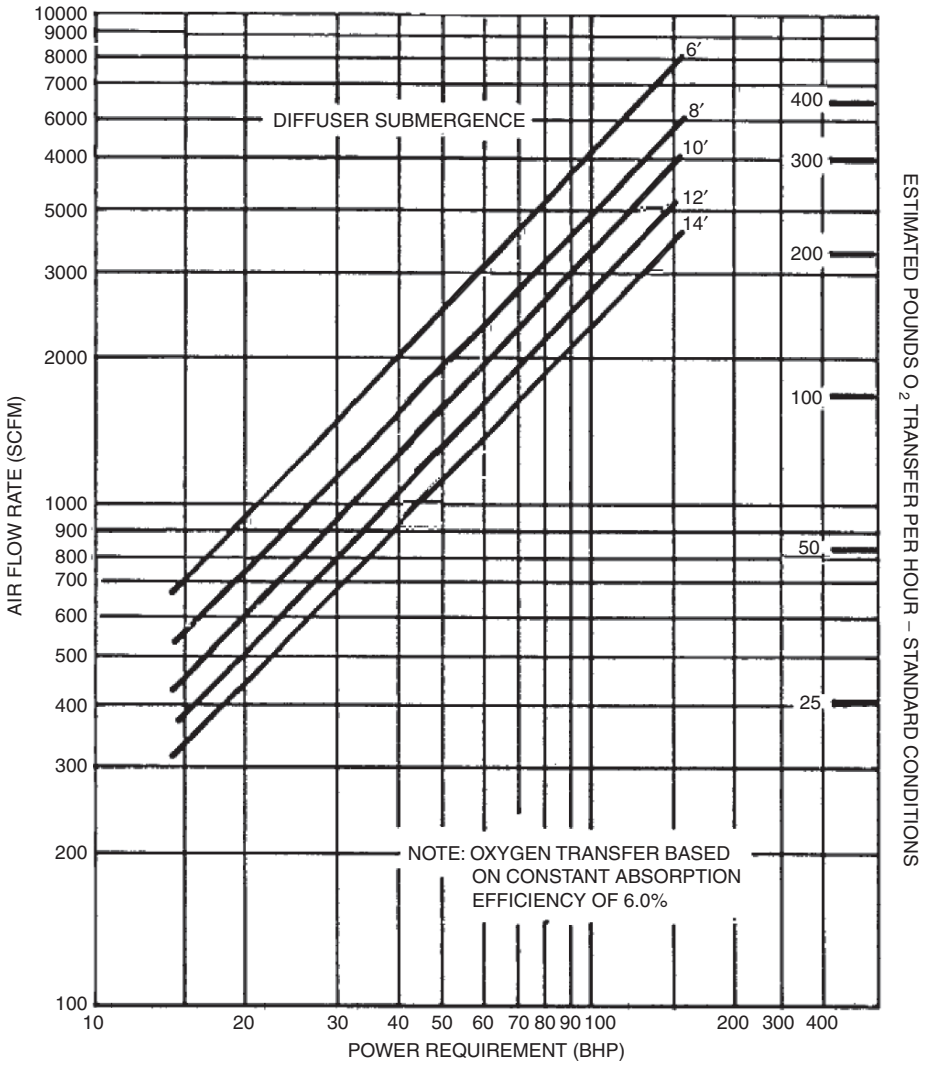

Fig. 4.4. Estimated horsepower requirements for air diffuser.

The long shaft is hard on bearings and is a source of maintenance problems. Submerged turbines have high maintenance requirements.

The sparge turbine aerator can be installed to augment existing diffused air systems. This offers such plants an opportunity to increase oxygen transfer capability with a moderate additional capital investment.

3.2.3. Static Mixing Aerator

The static mixer consists of a number of short elements of right- or left-hand helices. These elements are alternated and oriented so that each leading edge is at a 90° angle to the tailing edge of the one ahead. The element assembly is enclosed within a tubular housing.

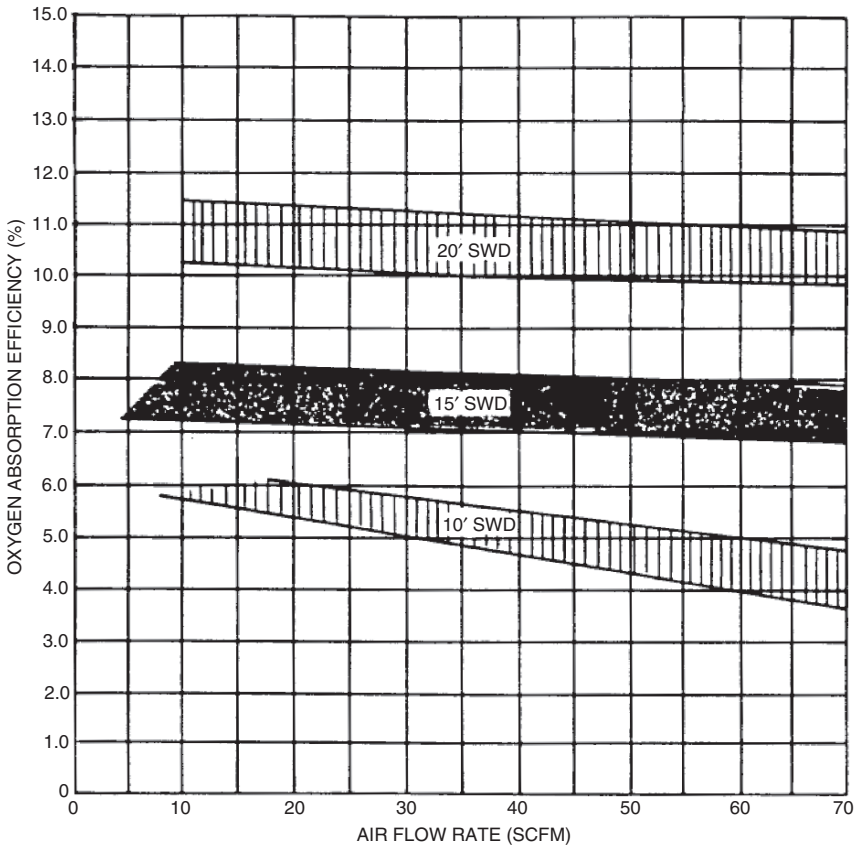

Fig. 4.5. Oxygen absorption efficiency vs air flow rate at various submergences of the static mixer.

Air is released beneath the mixers oriented vertically within the basin. The rising air contacts and entrains liquid as it rises through the mixer. The combination of large surface area and intense turbulence enhances oxygen transfer. Also the path of the air-water mixture is lengthened because of the winding channel imposed by the mixer units.

Figure 4.5 shows the oxygen absorption efficiency as a function of air flow rate at various submergences of the static mixer. Similar to air diffusers, the absorption efficiency of the static mixing aerator increases as the submergence increases for a fixed air flow rate. The static mixer also exhibits a slightly decreasing absorption efficiency as the air flow rate per tube increases. As shown in Figure 4.6, the oxygen transfer efficiency in terms of pounds oxygen transferred per horsepower exhibits a pattern similar to absorption efficiency. The transfer efficiency increases as submergence increases for a particular air flow rate.

3.2.4. Jet Aerator

The jet aerator uses apparatus for directing a stream or jet of air/liquid into the aeration basin to increase the oxygen content as well as liquid mixing. The pump recirculates the

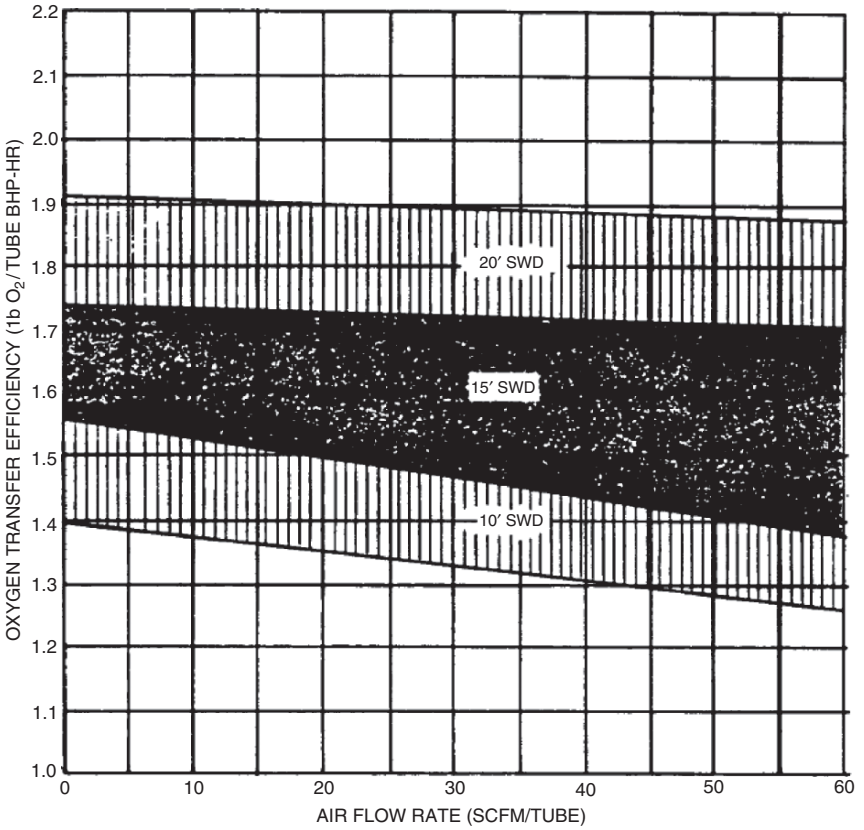

Fig. 4.6. Oxygen transfer efficiency vs air flow at various submergences of the static mixer.

liquid—a mixed liquor or a mixture of returned sludge, influent wastewater, and mixed liquor—through the jet aerator, to which air is introduced. The pressurized liquid shears off the compressed air into minute air bubbles.

Figure 4.7 shows typical performance curves for a jet aerator system. The performance curves show the dependence of oxygen transfer rate, absorption efficiency, transfer efficiency, and total power required as air flow rate varies. For example, as the air flow rate increases from 2100 to 3000 scfm (33.98 to 84.96 m^3/min), the oxygen transfer rate increases from 650 to 690 lb/h (294.1 to 312.2 kg/h); the absorption efficiency decreases exponentially from 36% to 25%; the total power required increases from 180 to 240 bhp (134.3 to 179 kw); and the oxygen transfer efficiency decreases slightly from 3.8 to 3.2 lb O_2/hp-h (2.83 to 2.39 kw-h). The performance of a jet aeration system is the result of interaction of several pertinent variables including jet submergence, ratio of air blower power to liquid pump power, air flow rate per jet, mixing cell diameter, liquid pumping rate per jet, and system oxygen demand.

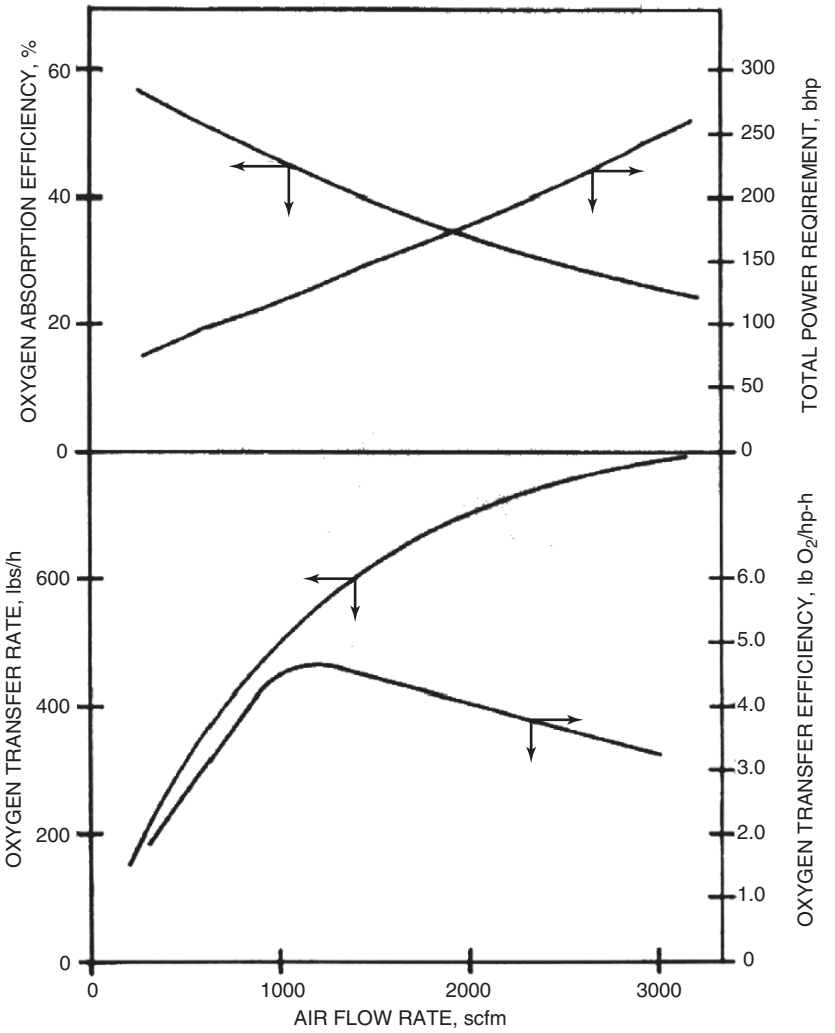

Fig. 4.7. Jet aerator performance characteristics as functions of air flow rate.

4. DESIGN APPLICATIONS

4.1. Types of Design Problems

In general, the design of a submerged aeration system can be categorized into three types:

1. To meet a certain oxygen demand.
2. To achieve the maximum oxygen transfer efficiency.
3. To achieve the maximum oxygen absorption efficiency.

In designing a new wastewater treatment plant that requires aeration processes, the oxygen demand is determined by the flow rate, strength of waste, characteristic of the waste, desired effluent quality and other factors. The submerged aeration system is designed to meet the maximum oxygen demand, which occurs at peak loading conditions, and, at normal conditions, the aeration system is operated to meet an average oxygen demand.

In the case where the power supply source is limited and the power consideration is a prime factor, the submerged aeration system is designed primarily to achieve the maximum oxygen transfer efficiency, i.e., minimum power required to meet the oxygen demand. The aeration system generally has the flexibility incorporated into the design such that the system will operate at the maximum oxygen transfer efficiency, while meeting the designed oxygen demand. For example, the air flow rate in a submerged aeration system can be designed and operated at a value such that the performance will be at maximum oxygen transfer efficiency.

In upgrading an existing treatment plant, an aeration system having maximum oxygen absorption efficiency is generally required because the existing air flow capacity is limited. In this case, the existing air supply system is used to increase oxygen transfer rate by changing or adding a new aeration system that has higher absorption efficiency than the existing aerator in terms of percent of oxygen transfer into the liquid.

Because aeration systems vary according to the nature of the products supplied by various manufacturers, and often proprietary information is involved in detailed design of an aeration system, an engineer usually consults manufacturers for a detailed design of a specific type of aeration system. In the following, a relatively new submerged aeration system will be covered as a case study.

4.2. Case Study Example

The case study was a biological extended aeration process treating pulp and paper wastewater using a submerged jet aeration system (32).

The design parameters of the plant were:

BOD loading	8500 lb/d (3855 kg/d)
Hydraulic loading	3 MGD (11,352 m^3/d)
Aeration volume	2.2 MG (8328 m^3)
BOD loading rate	28.9 lb BOD/1000 ft^3/d (0.46 kg/m^2/d)
Bottom maximum liquid velocity	1 ft/s (0.30 m/s)
Clarifier side water depth	17.5 ft (5.33 m)
Clarifier overflow rate	550 gpd/ft^2 (22.41 m^3/d/m^2)

4.2.1. Aeration System and Hydraulic Characteristics of the Channel

In striving for the most compact and economical design of an aeration basin, unique integrated clarifier-aeration basin was adopted. The aeration channel was located concentric with the final clarifier as shown in Figure 4.8. The clarifier inside diameter was 86 ft (26.2 m) with a 17.5 ft (5.33 m) water depth. The aeration channel was 37.25 ft (11.35 m) wide with a 20.5 ft (6.25 m) water depth.

The aeration equipment consisted of seven equally spaced directional mixed-jet manifolds, each with eight jets, from which all jets were mounted facing in the same direction,

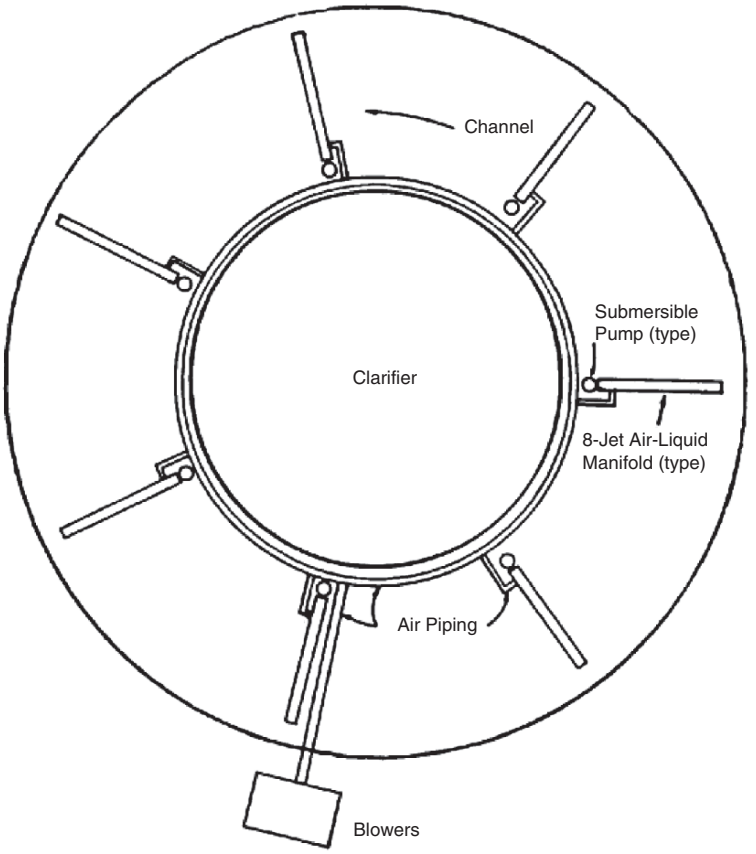

Fig. 4.8. Channel-clarifier jet manifold arrangement.

three centrifugal air blowers, and seven submersible sewage pumps. Each of the seven individual pumps was designed to provide motive fluid to all the jet aerators on a single manifold. Each pump-jet manifold combination formed a module aeration unit whose operation was not affected by the other six units. Each pump and aerator manifold was removable and replaceable in the basin without disturbing the operation of the other units.

The three centrifugal air blowers were each rated at 1250 scfm (35.40 m^3/min) and powered by a 75 hp (55.9 kw) motor. Two blowers were designed for use under normal conditions, and the third blower was provided as a standby unit. The blowers were located in a control building adjacent to the aeration channel. Air piping transported air from the blower across the aeration channel to the ring manifold located on the outside wall of the center clarifier. Seven individual air drop pipes were tapped into the ring manifold to direct air to the individual jet manifold. Butterfly valves were installed to control the air flow in the drop pipes. Each jet required a fluid flow rate of 90 to 100 gpm (0.34 to 0.38 m^3/min) at relatively low head conditions, 15 to

20 ft (4.57 to 6.10 m) water head. The air requirement was estimated at 40 scfm (1.13 m^3/min) per jet and the air pressure required was only slightly higher than the hydrostatic pressure exerted upon the jet orifice.

From this design data and the open channel hydraulics data (33, 34), an estimate of the water circulating velocity was made on the basis of water in the channel. The momentum input per jet was

$$\Delta M = \frac{Q\gamma_m U}{g} \tag{16}$$

where ΔM = momentum increase by one jet, lb; Q = air-water flow, cfs; γ_m = air-water mixture density, lb/ft^3; U = jet plume velocity right at the jet opening, ft/s; g = gravitational acceleration, ft/s^2. According to the momentum principle, the increase in elevation of free surface, Y, is:

$$\Delta M = \frac{Q\gamma_m Y J}{g\ A\gamma_w} \tag{17}$$

where Y = total elevation increase, ft; J = number of jets; A = channel cross-sectional area, ft^2; γ_w = water density, 62.4 lb/ft^3.

The increased water head is available to overcome friction and bend losses. The frictional loss, H_f, is estimated from Manning's equations:

$$H_f = \frac{n^2 V^2 L}{(1.49)^2 R^{4/3}} \tag{18}$$

where H_f = frictional loss along straight channel, ft; n = Manning's coefficient; V = liquid velocity in the channel, ft/s; R = hydraulic radius, ft and L = length of channel.

The bend loss, H_b, is estimated from data by Shukry (33):

$$H_b = f_c \frac{V^2}{2g} \tag{19}$$

where H_b = channel bend loss, ft; f_c = coefficient of curve resistance, which is taken as unit for the full 360° turn of the channel.

Equating the losses to the available head yields the velocity estimate:

$$Y = \left[\frac{n^2 L}{(1.49)^2 R^{4/3}} + \frac{f_c}{2g}\right] V^2 \tag{20}$$

The velocity estimated for this channel was 0.95 ft/s (0.29 m/s). Figure 4.9 shows a typical measured velocity profile inside the aeration channel. The velocity data were taken with a Savonius rotor current meter and the average velocity values are area weight average. The channel velocity was surveyed at an MLSS (mixed liquor suspended solids) concentration of 5300 mg/L. Two air flow rates checked were 20 and 40 scfm (0.566 to 1.13 m^3/min) per jet and the channel velocity measured were 1.06 and 0.95 ft/s (0.32 and 0.29 m/s), respectively, area weighted, which conformed closely to the estimated values.

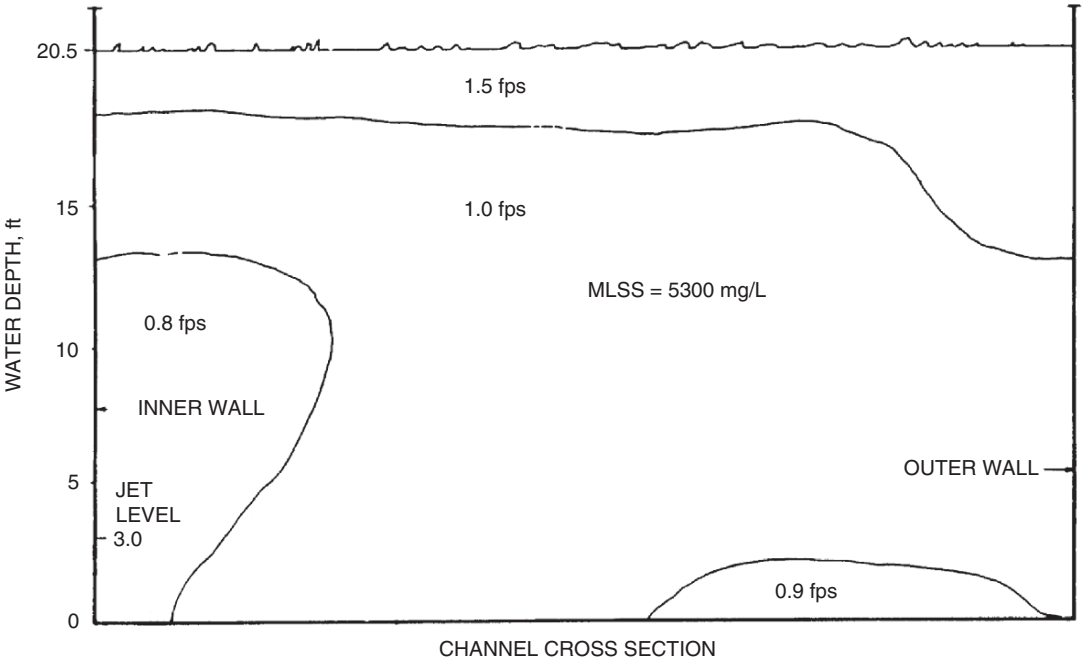

Fig. 4.9. Measured velocity profile within the channel.

4.2.2. Oxygen Transfer Efficiency Evaluation

A nonsteady states clean water oxygen transfer evaluation was conducted. Approximately 2200 lb (997.7 kg) of sodium sulfite and one pound (0.454 kg) of cobalt chloride were added. During the deoxygenation period, the seven pumps were turned on to continuously provide liquid mixing and maintain channel momentum.

Having the dissolved oxygen depleted to approximately zero, two of the three air blowers were turned on providing oxygen to the liquid. The oxygen concentration was determined simultaneously with two DO probes and samples taken and titrated by the Winkler method. The DO probes were standardized to the Winkler determinations. The oxygen saturation concentration, C_{ST} was determined at the end of each test by long-term aeration to reach the saturation equilibrium.

The power measurements were made with watt, volt, and ampere meters at the control panels. Brake horsepowers were calculated from the electrical parameter measurements. No corrections were made for electrical distribution losses.

Figure 4.10 shows the oxygen transfer efficiency at various air flow rates per jet. There existed an air flow rate at which the oxygen transfer efficiency was optimum. As shown in this figure, the highest oxygen transfer efficiency was about 4.2 lb O_2/hp-h (2.55 kg O_2/kw-h) at a flowrate of about 20 scfm (0.566 m^3/min) per jet. Applying air flow rates, other than 20 scfm per jet, result in lower oxygen transfer efficiencies. A significant drop in transfer efficiency was observed as air flow rate per jet increases.

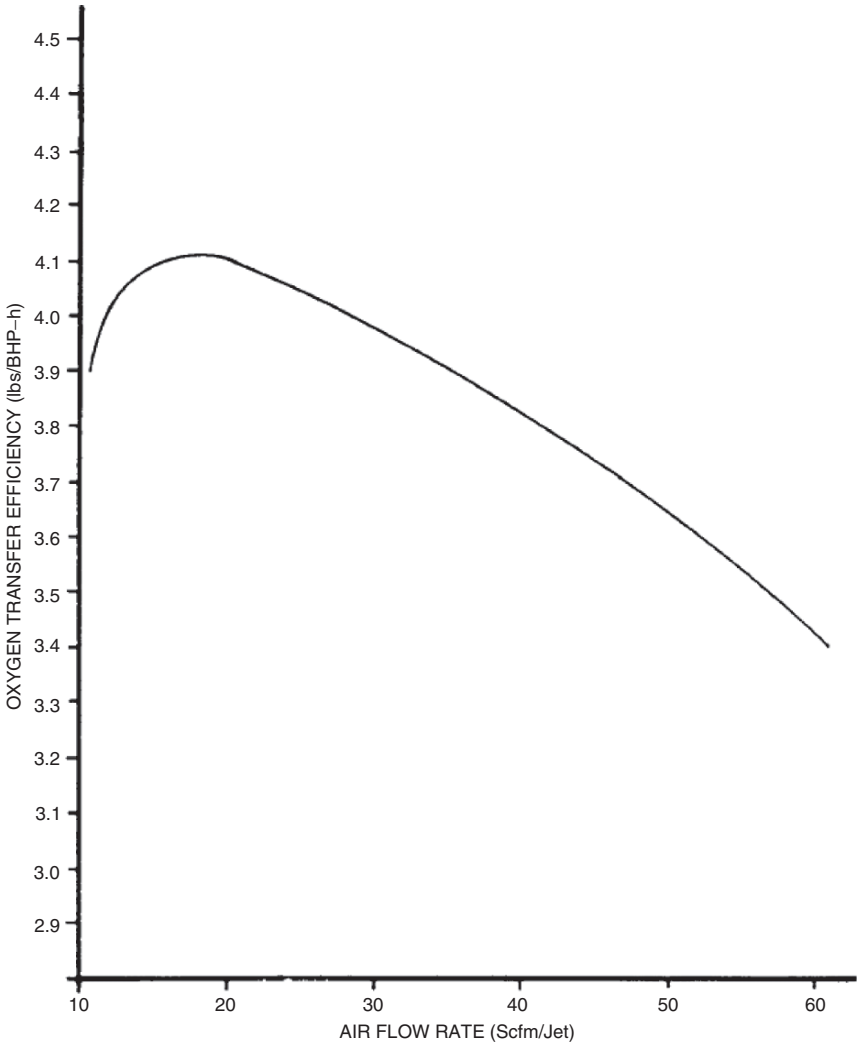

Fig. 4.10. Oxygen transfer efficiency vs air flow rate.

4.2.3. Operation and Results

The biological treatment plant was installed to treat the whole black liquor from both kraft and neutral sulfite pulp cooks, including the rinse water. The kraft cooks contain 5500 lb BOD_5 (2494 kg BOD_5)/d and the sulfite cooks contain 2450 lb BOD_5 (1111 kg BOD_5)/d. These cooking liquors are neutralized and then oxidized before being discharged to the aeration channel. The effluent from a business paper mill is used as dilution water for the cooking liquor. The paper mill manufactures coated grades and uncoated grades. This effluent contains up to 3000 lb BOD_5/d (1361 kg BOD_5/d) as well as significant amounts of clay from the coating.

The paper mill effluent is passed through a stationary bar screen and then pumped to the channel. The pulp and the paper mill flows are combined immediately ahead of the point of introduction into the channel. There is no primary sedimentation before aeration of wastewater.

The BOD_5 loading, expected to be 8500 lb/d (3855 kg/d), has averaged 11,000 to 12,000 lb/d (4989 to 5442 kg/d) instead. Actual BOD loading fluctuates over a wide range. Final effluent ranges between 7 and 21 mg/L BOD_5, and suspended solids content averages 26 mg/L.

This extended aeration system has overcome overloading, caustic shock, extreme variation in loading, inexperienced operation, high clay loading, prolonged high pH loading, high rate BOD loading, and winter 1emperatures. The cost for this 50,000 population-equivalent plant, including equipment and the construction costs for both aeration systems and final clarifier was $1,665,000 (1983 dollars).

5. RECENT DEVELOPMENT IN SUBMERGED AERATION

Researchers in recent years have directed their efforts to understand the relationship of specific air flow, submergence, diffuser density, bubble size, volumetric energy input and coverage mode (grid or floor) to the effectiveness bubble diffuser systems (35–38). Other researchers (38, 39) examined the effectiveness of jet aerator system and submerged stirring system. To save energy researchers (40, 41) have examined a method to determine when fine bubble aeration systems are operating efficiently and a method to control flow rates to fine bubble aeration systems. Cost saving method was presented by Wagner et al. (42) for clean water testing of fine bubble aeration system. Researchers (43–49) examined the effectiveness of submerged aeration system in individual household and small community biological treatment systems.

A paper presented by Wagner and Pöpel (35) examined the influence of the depth of submergence and the diffuser density on the volumetric oxygen transfer rate, the specific oxygen absorption and the aeration efficiency. The authors reviewed literature review on clean water oxygen transfer test data for aeration tanks with depth ranging from 3.5 to 12 m. It was determined that specific oxygen absorption is reduced at higher specific air flow rates (m^3/m^3-h at standard temperature and pressure) and greater depth of submergence. From the reported data, the authors determined that diffuser density in the range of 15% to 20% had specific oxygen absorption of approximately 20 g O_2/m^3-m^2) at specific air flow rate of 2 m^3/m^3-h. When the diffuser density was lower to 5% to 10%, the specific oxygen absorption was decreased to approximately 13 g O_2/m^3-m^2 at specific air flow rate of 2 m^3/m^3-h. It was also observed that specific oxygen absorption decreased when the specific air flow rate increased. The author showed that high specific oxygen absorption is achieved for fine bubble diffusers in clean water with shallow tanks (3.5 to 6.0 m), high diffuser density (25% to 20%) and at low specific air flow rates (less than 3 m^3/m^3-h).

Höfken et al. (36) presented a discussion on fine bubble membrane diffusers, which were composed of elastic perforated material. Some of commonly used materials used today are synthetic rubber and polyurethane. These materials are perforated by punching holes or slits.

The authors reported that the quality assessment of aeration system can only differ in their range of bubble sizes generated by the aeration diffusers and the energy used for the bubble generation. It is suggested by the authors that the maximum oxygen input is achieved with gas bubbles diameters ranging from 1.3 to 2.3 mm. Silicon fine bubble diffusers were examined by the authors. These diffuses are perforated flexible hose with both ends of tube attached to the air distribution pipe and the tubing is mounted to the bottom of the basin. To prevent excess curving of the silicon tubing, intermediate clamping bar are provided to hold down the tubes. The upward flow and the curving of the tubing results in an oscillating movement that results in higher bubble separation frequencies and prevents depositions forming under the diffusers. These silicone membranes are reported to have gas bubbles diameter ranging from 1.3 to 2.3 mm and achieve oxygen yields ranging from 4 to 6 kg O_2/kw-h.

Newbury (37) examined performance database of fine bubble aeration system operating in predominantly floor coverage and grid type modes. The data included diffuser type, diffuser size, submergence, airflow per diffuser, the ratio of tank area per diffuser area and clean water standard oxygen transfer efficiency (SOTE). The diffuser types were perforated membrane sheets, membrane discs, membrane tubes, ceramic domes, ceramic domes, impacts and plastic discs. Tables 4.2 and 4.3 present the database of fine bubble diffusers examined by author.

Table 4.2
Fine bubble diffusers database: size/diffuser, submergence and air/diffuser

Type/Coverage	Size/Diffuser (ft^2)	Submergence (ft)	Airflow/Diffuser (Std ft^3/sec)
Perforated Membrane Sheet/Floor	40	14.3, 15.5, 16.0 & 29.3	6.66E-02–6.0E-01
Membrane Disc/Grid	0.267, 0.442 & 0.785	2.7, 15.8–22.8 & 29.2	2.41E-04–1.92E-03
Membrane Tube/Grid	0.243	12.8, 13.5, 17.5 & 19.0	2.07E-02–5.47E-02
Ceramic Dome/Grid	0.267	14.0 & 15.0	1.90E-02–4.15E-02
Ceramic Disc/Grid	0.267	14.4, 14.8, 15.8, 17.0 & 18.7	1.20E-02–4.8E-02
Plastic Disc/Grid	0.267	13.0	6.93E-02
Impact	0.349	2.7	8.83E-03–2.15E-02

Table 4.3
Fine bubble diffusers database: number of tests, tank area/diffuser area and SOTE

Type/Coverage	Number of Tests	Tank Area/Diffuser Area (ft^2/ft^2)	Clean Water SOTE (%)
Perforated Membrane Sheet/Floor	29	1.58–3.91	29.43–59.74
Membrane Disc/Grid	61	4.80–20.1, 40.18 & 49.60	5.50–56.95
Membrane Tube/Grid	11	16.1–60.3	11.92–32.79
Ceramic Dome/Grid	22	10.81–32.9	6.50–29.49
Ceramic Disc/Grid	10	13.7–16.1	25.29–36.83
Plastic Disc/Grid	1	21.3	23.21
Impact	4	10.81	4.60–5.90

Using this database, the author developed a relationship between volumetric energy input or energy intensity (E_I; kw/m^3) for a fine bubble aeration device and the standard oxygen transfer rate (SOTR, g/m^3-sec). Author proposed the following:

$$\text{SOTR} = -11.5{E_I}^2 + 2.27E_I \tag{21}$$

Newbury suggests that this relationship between SOTR and E_I provide for improving oxygen transfer efficiency for fine bubble aeration system. This can be accomplished by reducing the bubble sizes or by reducing the energy applied to the fine bubble aeration system. The rate of oxygen transfer per unit volume of clean water is proportional to uniform diameter (D) of the bubbles raised to negative seven fourths ($D^{-7/4}$).

Lane (38) presented a discussion on both fine bubble and course bubble diffuser systems in lagoon. A general rule of thumb, that course bubble diffusers have OTE of 0.75% per foot of depth of submergence and fine bubble diffusers have OTE of 2% per foot of depth of submergence. The author discussed static tubes, coarse bubble diffuser system that was typically used in lagoons. These tubes are installed on the bottom of the lagoon with air traveling several feet up tube before discharging air into the upper layers of the lagoon. As result of this design, inadequate aeration and mixing are common problems for lagoon with static tube, because the lower portions of the lagoon is not receiving sufficient aeration or mixing. As an improvement, coarse bubble diffusers system are designed to release the near the bottom to ensure adequate aeration and mixing. The alpha factor for diffuser equipment provides a method for adjust clean water test result that manufacturer provide for their equipment to wastewater application. For both coarse bubble and fine bubble system, the alpha factor range from 0.5 to 0.8.

The alpha factor for jet aerator system was compared to member diffuser system by Backman et al. (39). Because the alpha is directly proportional to the SOTR calculated for design, the higher the alpha value for equipment in a wastewater the higher the oxygen transfer rate, which has major influence on capital cost and power consumption aeration diffuser systems. The results of pilot study, conducted on an activated sludge process treating wastewater from several pulp/paper mills, showed the alpha values ranged from 1.78 to 3.18 for jet aerators and 0.53 to 0.70 for membrane diffuser system.

Höfken et al. (36) also discussed the velocity required to suspend activated sludge in an aeration tank. For submerged stirring systems in circular and rectangular tanks, the most favorable energy condition for suspension is obtained by centrally locating the bottom stir system. To determine the minimum tank bottom velocity consideration must be given to the locations in the tank that have high potential for solids to settle. These locations include the edges of the tank and the point where the horizontal and vertical flows meet. Additionally, it is also necessary to determine the equivalent weight of the particles and match the roll resistance caused by the flow forces acting on the particle. For typical particle size of 60 μm, a bottom velocity of 10 cm/s is sufficient to maintain activated sludge in suspension, while for particle size as large as 100 μm a bottom velocity of 15 cm/s is sufficient to maintain suspension. The authors examined a bottom hyperboloid stirring aeration system. This system includes a hyperboloid stirrer with transport ribs attached to upper surface of the stirrer. The shape of the stirrer induces a flow that remains attached to the upper surface of the stirrer and ribs

transport the fluid away from the stirrer which result in the mixing of the tank contents. Air is introduced using sparge ring beneath the stirrer. Air bubbles are distributed throughout the basin by the macro-motion induced by the transport ribs. Authors recommend the use of the hyperboloid stirring system for anoxic tanks. The power consumption in the denitrification mode for this mixing system ranged from 1 to 2 w/m^3. The hyperboloid stirring with aeration system is recommended for intermittent aeration needs or nitrification tanks. Depending on the shape of the reactor and oxygen demands, the hyperboloid stirring with aeration system has oxygen yields of 2.0 to 2.9 kg of oxygen/kw-h.

Iranpour et al. (40) examined the air flux ($Nm^3/min\text{-}m^2$) and oxygen transfer efficiency (OTE) for a fine bubble diffuser system in aeration basins at Tillman Wastewater Reclamation Plant in San Fernando Valley, California. The author showed that by monitoring the air flux throughout the length of an aeration basin that a drop in OTE (%) can be predicted by increase in air flux. It suggested that this increase in air flux can be used to predict when air diffuser system requires maintenance, which includes cleaning diffusers, repair leaking air piping or replacing diffusers.

Ferrer et al. (41) conducted a bench-scale experiment on a Bardenpho process to compare the on-off aeration control system to a fuzzy logic based aeration control system. The fuzzy control system makes it possible to build controllers based on inexact or fuzzy information that may be obtained directly from plant operators. The bench-scale process was fed with wastewater that was similar in characteristics to a typical municipal wastewater. Like most typical wastewater both the hydraulic and organic loadings fluctuated. The researchers' goal was to maintain a constant 2.5 mg/L of dissolved oxygen level in the aeration basin as demand changed owing to the fluctuation in hydraulic and organic loadings. By maintaining a constant dissolved oxygen level in the biological reactors, the biological process would be more efficient, which would result in less energy consumption. Solenoid valves were used to control the air flow rate to biological process. The control elements were wired through a data acquisition card to a PC where the off-on and fuzzy logic control systems were implemented. The experimental data indicated that the dissolved oxygen level in the biological reactor ranged from 2.2 to 2.9 mg/L for the off-on control system and ranged from 2.4 to 2.6 mg/L with the fuzzy logic control system. A 20% or more energy saving could be realized by using a fuzzy logic aeration control system over an off-on aeration control system.

To guarantee performance of fine bubble aeration diffuser system, designers are requiring fine bubble aeration diffuser system to be tested in clean water before start-up phase of wastewater treatment plant. The testing determines if the fine bubble aeration diffuser system is meeting the oxygen transfer rate requirements. Wagner et al. (42) compared a newly proposed pure oxygen desorption method to the unsteady absorption method. The latter method is standardized in Germany and Austria. The unsteady absorption method was published in 1993 in the United States by the America Society of Civil Engineer (ASCE) in the Standard: Measurement of Oxygen Transfer in Clean Water. This method requires the addition of sodium sulfite and cobalt salt as a catalyst of the sulfite oxidation. The addition of the sulfide lowers the oxygen concentration in the clean water to zero. The fine bubble aeration diffuser system is activated and the oxygen transfer curve from zero to saturation is used to determine the aeration coefficient $K_L a$ and the oxygen saturation concentration. A problem with this method

includes the uniform distribution of the sulfide into the water which ensures that oxygen level through the tank is zero. As an alterative to this method Darmstadt University of Technology developed the pure oxygen desorption method.

Supplementing the air to the fine bubble diffuser system with pure oxygen increases the oxygen saturation concentration in the clean water to a point above the saturation point. As a result, the reduction of excess oxygen over a period of time down to saturation point is calculated by the following equation.

$$c_{(t)} = c^* - (c^* - c_0)e^{-K_L a \cdot t} \tag{22}$$

where:

c^* = oxygen saturation concentration under process condition, g/m^3
c_0 = oxygen concentration at time $t = 0$, g/m^3
K_La = aeration coefficient, 1/h
t = time, min

Equation (22) is the desorption equation, which describes the reduction of excess oxygen concentration with time to the saturation concentration. Because this equation is the same equation that describes the increase of oxygen level from zero to saturation concentration, the resulting aeration coefficients are equal.

To use the pure oxygen desorption method, the pure oxygen flow rate, dosing time for pure oxygen in the air header pipe and the amount of pure oxygen required. The pure oxygen flow rate Q_{O2} is determined as follows:

$$Q_{O2} = Q_a([21c_{s'}/c_s] - 21)/(100 - [21\ c_{s'}/c_s]) \tag{23}$$

where:

$c_{s'}$ = oxygen concentration in the air/oxygen mixture, g/m^3
Q_{O2} = pure oxygen flow rate, m^3/h at STP
Q_a = air flow rate, m^3/h at STP

To prevent explosion, the oxygen concentration ($c_{s'}$) must be smaller than twice the oxygen saturation concentration (c_s). The required dosing time for the pure oxygen into the air header is determined in Equation (24). The excess saturation concentration in the water, Δc, is selected so that difference is greater than 6 g/m^3. This excess saturation concentration is also dependent on length of pipe in the air piping system and dosing location for the pure oxygen. When the air system is long the Δc valve is set higher than for shorter air pipe system. Equation (24) is used to determine the dosing time for in the air header.

$$t_{O2} = (60/K_L a') \ln[(c_{s'} - \Delta c - c_s)/(c_{s'} - c_s)] \tag{24}$$

where:

Δc = excess oxygen concentration, g/m^3
t_{O2} = dosing time in the header, min
$K_L a'$ = aeration coefficient for the air/oxygen mixture, 1/min

The amount of pure oxygen required is determined by using Equations (25) and (26).

$$M_{O2} = (Q_{O2}t_{O2})/60 \text{ (m}^3 \text{ at STP)} \quad (25)$$

$$M_{O2} = (Q_{O2}t_{O2})(1429)/60 \text{ (kg)} \quad (26)$$

The authors compared the absorption method to the pure oxygen desorption method in a race track aeration tank at the Darmstadt wastewater treatment plant in Germany. The aeration tank volume has a volume of $18,000\,m^3$ and equipped with fine bubble membrane tube diffusers. Under the absorption method, it required 11,600 kg of sodium sulfite and 37 kg of cobalt chloride as catalyst to remove the oxygen from the water. The total chemical cost for the absorption method was reported to be approximately twice as much as the pure oxygen desorption method. The authors also reported that absorption method is difficult because distribution of sodium sulfite with the cobalt chloride is time consuming and hard to complete when compare to the pure oxygen desorption method.

Daude and Stephenson (43) examined a newly developed package plant that treats domestic wastewater from single household. The package plant combined submerged aerated filter (SAF) with jet aeration into a compact shadow tank. The authors concluded that jet aeration proved to be the best aeration method for shallow bioreactor design with removal efficiencies of 94.2% for BOD_5, 85.9 for COD and 87.6% for SS. The package plant produced inconsistent values for effluent ammonia varying from 9 to 60 mg/L. Additional technical information on SAF can be found from the literature (47–49). Myung and Yu (44) examine the retrofitting of sequential batch reactor with an intermittent aeration system in small community. The evaluation indicated that the organic and nitrogen removals were accomplished with the intermittent aeration, but phosphorus concentration increased. The authors concluded that the increase was owing to release of phosphorus under the nonaeration periods.

Wastewater treatment facility (24.1 MGD maximum month) in SW Florida used coarse bubble diffused aeration system in two sludge holding tanks (45). The tanks receive waste activated sludge (WAS) with TSS solids concentration ranging from 0.8% to 1.0%. Diffused aeration system was design first to maintain solids in suspension and second to keep the stored WAS aerobic before dewatering by belt filter presses. Tanks were designed in a fill and draw operational mode. The dimensions of the tanks are:

Length, ft	149.75
Width, ft	30
Max Side Water Depth, ft	16.34
Min Side Water Depth, ft	2
Floor Area, ft^2	4493
Max Volume, ft^3	73407
Min Volume, ft^3	8985

Wilfey Weber, Inc. provided Model CBS-24 stainless steel diffusers with $5/8''$ orifices that have a $24''$ long deflector with 6-0.188 and 6-0.38″ diameter holes (45).

The diffuser system had following criteria:

Air required @ Max Depth	1900 scfm/basin
Air required @ Min Depth	500 scfm/basin
Mixing Energy@ Max Depth	25.9 scfm/1000 ft^3
Mixing Energy@ Min Depth	55.6 scfm/1000 ft^3
Number of diffusers	84/basin
Max Air/diffuser	22.62 scfm
Min Air Flow/Diffuser	5.95 scfm
SOTE @ Max Depth	15.65%
SOTE @ Min Depth	0.727%
Max Standard Oxygen Delivered	310.2 lbs O_2/h
Min Standard Oxygen Delivered	3.8 lbs O_2/h
Diffuser Headloss	0.22021 psi

Because the water level in the tank will vary, positive displacement blower were provided for each tank with one backup blower.

Fine bubbles bubble diffused aeration system (Sanitaire) was installed at South County Water Reclamation Facility (16.0 MGD maximum month), Collier County Florida (46). The diffusers were 9 inch diameter disc that were designed to meet oxygen requirements for carbonaceous removal and nitrification in Modified Ludzack-Ettinger (MLE) activated sludge process.

The dimensions of the tanks and diffuser system requirement are listed below:

Length, ft	127.33
Width, ft	17.0
Max Side Water Depth, ft	20.40
Min air flow, scfm/Flow per diffuser scfm/diffuser	470/1.01
Avg air flow, scfm/Flow per diffuser scfm/diffuser	1050/2.26
Max air flow, scfm/Flow per diffuser scfm/diffuser	1400/3.01
SOTE at minimum air flow, %	37
SOTE at average air flow, %	34.1
SOTE at maximum air flow, %	32.9
Diffuser Orifice Headloss with one $13/64''$ port, psi	4.04×10^{-2}

To meet the air flow range noted above, four multi-stage centrifugal blowers were provided: two 7-stage blowers at 4500 scfm/each and two 8-stage blowers at 1950 scfm/each.

NOMENCLATURE

A = channel cross sectional area, L^2
$c_{s'}$ = oxygen concentration in the air/oxygen mixture, ML^{-3}
Δc = excess oxygen concentration, ML^{-3}
c^* = oxygen saturation concentration under process condition, ML^{-3}
c_0 = oxygen concentration at time $t = 0$, ML^{-3}
C = dissolved oxygen concentration, ML^{-3}

C_o = dissolved oxygen concentration at $t = 0$, ML^{-3}
C_S = dissolved oxygen concentration in pure water of temperature T,
C_{SS} = disso1vcd oxygen concentration corrected to standard conditions, ML^{-3}
C_{ST} = average mid-depth oxygen saturation concentration, ML^{-3}
D_{inches} = diameter of pipe inches for air pipe, inches
DO_{760} = dissolved oxygen saturation value in distilled water of temperature 22° C under 1 atm, mg/L
e = oxygen absorption efficiency
E = oxygen transfer efficiency, $MLFT^{-2}$
E_I = volumetric energy input or energy intensity for a fine bubble aeration device, $MLFT^{-2}$
f_c = coefficient of curve resistance
F_i = mole fraction of oxygen in the exit air
F_o = mole fraction of oxygen in the feed air
g = gravitational acceleration, LT^{-2}
h_f = frictional loss in air piping, ft
H = Henry's constant, FL^{-2}
H_b = frictional loss owing to channel bend, L
H_f = frictional loss along straight channel, L
J = number of jets
K_La = overall oxygen transfer coefficient, T^{-1}
K_La' = aeration coefficient for the air/oxygen mixture, T^{-1}
M_{O2} = required amount of pure oxygen, L^{-3}
ΔM = momentum increase by one jet, F
n = Manning's coefficient
N_o = oxygen transfer rate under standard conditions (OTR), MT^{-1}
p = pressure of saturated water vapor at the temperature of the water, mm
P_a = one atmosphere pressure, FL^{-2}
P_A = actual pressure, FL^{-2}
P_B = atmospheric pressure, barometer, FL^{-2}
P_S = standard pressure, FL^{-2}
PV_A = actual vapor pressure of water at actual temperature, FL^{-2}
PV_S = standard vapor pressure of water at actual temperature, FL^{-2}
Q = air flow rate, L^3T^{-1}
Q_{O2} = pure oxygen flow rate, L^3/T at STP
Q_a = air flow rate, L^3/T at STP
Q_{acfm} = actual air flow rate at site conditions, cfm
Q_{scfm} = standard air flow rate at standard conditions, cfm
R = hydraulic radius, L
RH_A = actual relative humidity
RH_S = standard relative humidity
t = aeration time, T
t_{O2} = dosing time in the header, T
$SOTR$ = standard oxygen transfer rate, $ML^{-3}\ T^{-1}$

$SOTE$ = standard oxygen transfer efficiency, %
T = water temperature, degree
T_A = actual temperature, °R
T_S = Standard temperature, °R
U = jet plume velocity at the jet opening, LT^{-1}
V = water movement velocity in the channel, LT^{-1}
V_{fpm} = air velocity in pipe, fpm
W = mass of water being aerated, M
Y = increase in elevation of free surface L
Z_d = aerator submergence, L
γ = microbial oxygen uptake rate, $ML^{-1}T^{-1}$
γ_w = mass density of water, ML^{-3}
γ_m = mass density of air-water mixture, ML^{-3}
α = ratio of K_La in wastewater to K_La in clean water at equivalent conditions of temperature. mixing. etc.
β = ratio of dissolved oxygen saturation to that in clean water at equivalent conditions of temperature and partial pressure
θ = temperature correcting coefficient
φ = constant characteristic of the aeration bubble pattern

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5
Surface and Spray Aeration

Jerry R. Taricska, J. Paul Chen, Yung-Tse Hung, Lawrence K. Wang and Shuai-Wen Zou

CONTENTS

Abstract Surface aeration involves the use of special floating aerators or spray aerators for removing taste/odor-causing substances, carbon dioxide, hydrogen sulfide, methane, volatile organic compounds, etc. from water, and for oxidizing iron and manganese in drinking water. The topics covered in this chapter are: gas solubility, diffusion, equilibrium, mixing, gas transfer, reaeration, instream aeration, surface aeration, spray aeration, and engineering design. Also included are the design examples, and the case histories of hydrogen sulfide removal, deferrization, demanganization, taste and odor removal.

Key Words Surface aeration • spray aeration • gas transfer • aerators • reaeration • deoxygenation • mechanical aerator • instream aeration • nozzle aerator • multiple-tray aerator • hydrogen sulfide removal • carbon dioxide removal • deferrization • demanganization • taste and odor removal.

1. INTRODUCTION

Gas transfer in water and wastewater treatment is a process whereby water is brought into contact with air or a gas and, because of the presence of a concentration gradient, the transfer

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of gases to and/or from the water occurs. In water treatment, for example, aeration is used to remove carbon dioxide, hydrogen sulfide, methane, and various volatile organic compounds, and to oxidize iron and manganese in the water. In wastewater treatment, the gas transfer is used primarily in the activated sludge and trickling filtration processes and in aerated lagoons and aerobic digesters. Air or oxygen gas is supplied to the wastewater by air compressors and by mechanical aerators to induce a higher concentration gradient in the wastewater and thereby to accelerate oxygen transfer. The spray of wastewater by nozzles over trickling filter beds also results in an accelerated oxygen transfer. In polluted streams and lakes, oxygen transfer from the limitless atmosphere to the water occurs, also because of the existence of dissolved oxygen concentration (or deficit) gradient. This reaeration process is an important and essential element of the stream self-purification process.

Gas transfer is a transport phenomenon in which molecular diffusion plays a significant role. Molecular diffusion effects can be described as the product of molecular diffusivity and the concentration gradient, as defined by Fick's Law. The value of diffusivity is almost always determined experimentally, although it can be predicted theoretically. The determination of the gradient is a very complicated one and hence has received considerable study, theoretically and experimentally. The various theories generally developed on the assumption that there exist at the interface both gas and liquid films, can be classified into: (a) the film model, (b) surface renewal models, and (c) other models representing various fields of interest. This chapter describes the various models of gas transfer, and compares their gas-liquid mass transfer coefficients. Applications to surface and spray aerations and to natural reaeration are also presented.

2. FUNDAMENTAL CONCEPTS

Before the theory of gas transfer is presented, several important physical and chemical principles of gases and liquids, and the properties of the gas-liquid interface, are discussed in the following.

2.1. Equilibrium

In a quiescent body of water, the concentration of dissolved substance will eventually become uniform and no further change in concentration will occur with time. The concentration at this condition is called the equilibrium or saturation concentration. The equilibrium concentration for dissolved oxygen in water is 9.02 mg/L at 20°C water temperature and 1 atm pressure (760 mm Hg). Water exposed to air tends to reach this equilibrium. If the water is undersaturated with oxygen, oxygen will go into solution and equilibrium will gradually be approached from the direction of undersaturation toward saturation. If, on the other hand, the water contains more oxygen than the saturation quantity, oxygen will be released from the water and equilibrium will be approached from the direction of supersaturation. The former process is termed gas absorption and the latter, gas release or gas stripping.

The equilibrium or saturation concentration is an important concept because it is the difference between this equilibrium gas concentration and the actual gas concentration in the

Table 5.1
Absorption coefficients for gases in water and water vapor pressures

Temperature °C	Absorption coefficients[a]							Vapor pressure[c]
	Air[b]	N_2	O_2	CO_2	H_2S	CH_4	NH_3	
0	28.8	23.0	49.3	1710	4690	55.6	1300	4.58
10	22.6	18.5	38.4	1190	3520	41.8	910	9.21
20	18.7	15.5	31.4	878	2670	33.1	711	17.5
30	16.1	13.6	26.7	665	2120	27.6	—	23.8

The equilibrium concentration of oxygen in pure water of 0°C at 1 atm can be computed, by Eq. (1).

[a] The absorption coefficients k_s are given in milliliters of gas per liter of water at 760 mm Hg partial pressure of gas.

[b] Contents of a dry air by volume at sea level: 78.03% N_2. 20.99% O_2. 0.9323% argon. 0.03% CO_2, 0.00241% rare gases. 0.00006% O_3.

[c] Vapor pressure of pure water p_y, mm Hg. at atmospheric pressure.

liquid that provides the driving force (concentration gradient) to cause gas transfer into or out of the liquid. This equilibrium gas concentration is the solubility of the gas in the liquid.

2.2. Gas Solubility

When a liquid, such as water, and a gas are in equilibrium, the amounts of gas in solution may be determined by Henry's law. Henry's law states that for dilute solutions at equilibrium, the saturation concentration of a gas in liquid is directly proportional to the partial pressure of the gas in the vapor phase in contact with the liquid, i.e., equilibrium exists between the same gaseous constituent of liquid and that of the vapor. It is generally written as:

$$C_S = k_S p \tag{1}$$

where

C_S = saturation concentration of the gas in the liquid,
p = partial pressure of the gas in the vapor,
k_S = proportionality constant or the absorption coefficient.

The absorption coefficients for several gases are listed in Table 5.1.

Dalton's law of partial pressures states that the molecules of each gas in a gas mixture exert a partial pressure of each gas and the sum of these partial pressures equal the total pressure. Based on Avogadro's hypothesis that equal volumes of ideal gases contain, at the same temperature and pressure, equal numbers of molecules, the partial pressure of a gas in the atmosphere can therefore be calculated by proportion to the volume content of the gas in the air. Because the air in contact with water is generally saturated with water vapor, the vapor pressure of water must be subtracted from the total pressure in the calculation. The contents of dry air and the vapor pressure of water are also included in Table 5.1. The equilibrium

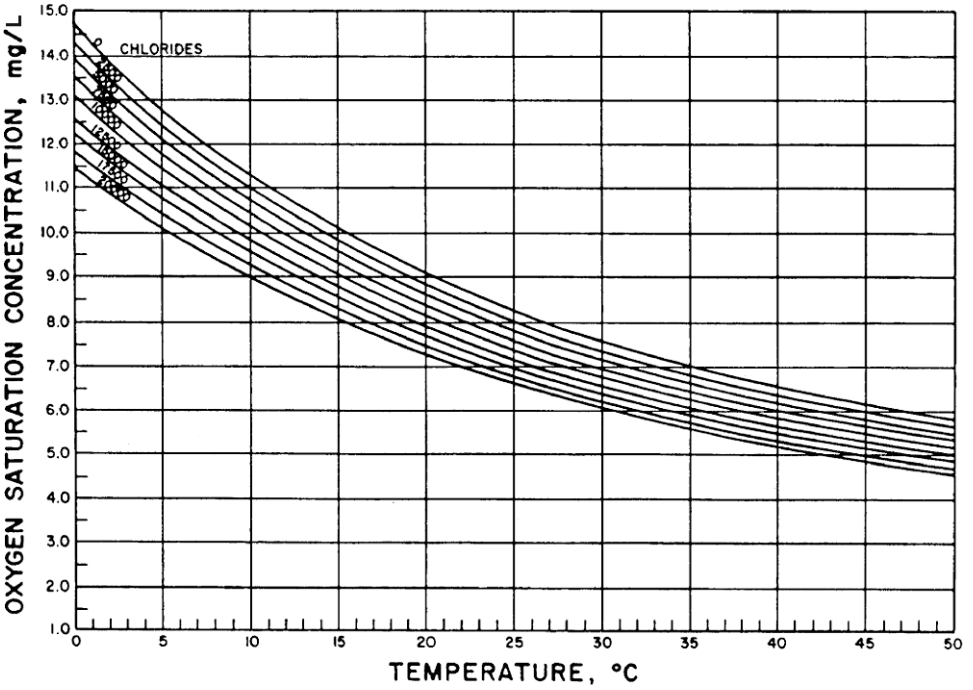

Fig. 5.1. Oxygen saturation concentration.

concentration of oxygen in pure water of 0°C at 1 atm can be computed, by Equation (1),

$$C_S(0°C) = 49.3 \times \frac{0.2099 \times (760 - 4.58)}{760}$$
$$= 10.28 \text{ oxygen mL/L} = 14.71 \text{ mg/L}$$

where, from Table 5.1, $k_s = 49.3$ mL/L at 760 mm Hg, the dry air includes 20.99% of oxygen by volume, the vapor pressure of water is 4.58 mm Hg, and 1 mL of oxygen weighs 1.43 mg.

Henry's law applies closely to gases that do not react chemically to any great extent with water. Oxygen, nitrogen, carbon dioxide, and the rare gases are of this kind. Other gases such as methane and hydrogen are inert; hydrogen sulfide less so. Chlorine and ammonia are strongly reactive, and readily form compounds with water. Gas solubility decreases with increasing temperature, as indicated by the decreasing values in the absorption coefficient for increasing temperatures. The oxygen saturation concentration in water at various temperatures may be computed by

$$C_S(T°C) = 14.61996 - 0.40420T + 0.00842T^2 - 0.00009T^3 \quad (2)$$

where:

T = temperature, °C

C_S = saturation concentration of the gas in the liquid, mg/L (3).

Dissolved solids such as salt lower the solubility of gases in water. Figure 5.1 provides curves for obtaining oxygen saturation concentration at various chloride concentration and temperature. The equilibrium concentration of dissolved oxygen in sea water (with 20,000 mg/L chloride concentration) at 1 atm pressure, 20°C, is 7.2 mg/L or about 80% that of fresh water. Hence, the solubility of oxygen reduces about 5% for every 5000 mg/L of chlorides. For domestic wastewaters the dissolved oxygen saturation is about 95% of that in fresh water. The oxygen saturation concentrations presented by the American Public Health Association (a) and the Committee on Sanitary Engineering Research (b) have generally been accepted as standards. Soubilities of other gases at various temperatures are also available elsewhere (1, 4, 5).

In summary, the solubility of a gas in liquid depends upon: (a) its partial pressure of the gas in contact with the liquid, (b) the liquid temperature, (c) the dissolved solids in the liquid, and (d) the chemical nature of the gas.

2.3. Molecular Diffusion

Molecular diffusion is a purely physical process. For example, if a few crystals of copper sulfate are added to water, the blue-colored copper sulfate solution would gradually spread out through the water and eventually will become uniform throughout the volume of water. This slow mixing of the copper sulfate with the water, without any movement of the water itself, is achieved as a result of molecular diffusion. All molecules possess their own inherent kinetic energy, which can be expressed in terms of mass and velocity. Molecules of specific mass m will move about with a specific velocity v, according to the relation:

$$\text{Kinetic energy} = \frac{1}{2} m v^2 \tag{3}$$

The movement of molecules is entirely random and it is this movement that allows the dissolved molecules to spread out or diffuse and eventually achieve a uniform concentration.

Fick's law of diffusion (6) provides a quantitative determination of the molecular diffusion. Fick's law states that the rate of mass transfer per unit area (i.e., mass flux) is proportional to the concentration gradient across the unit area, or

$$n = -D_{\mathrm{m}} \frac{\mathrm{d}C}{\mathrm{d}y} \tag{4}$$

where:

n = net mass flux across a unit area,
$\mathrm{d}C/\mathrm{d}y$ = concentration gradient across the unit area,
D_{m} = coefficient of molecular diffusion or the molecular diffusivity.

D_{m} is a function of the liquid temperature and hence related to the viscosity and density of the liquid, and the size of the diffusing molecules or the molecular weight.

As defined by Fick's law, the driving force for the diffusion is the concentration gradient of the substance being diffused. In completely quiescent oxygen deficient water, the concentration gradient is measured as vertical variation or $\Delta C/\Delta h$ where Δh is a depth increment and. ΔC is the concentration difference. The diffusion process involves the entry of oxygen

molecules from the air into the water at the air-water interface and subsequent distribution of the dissolved molecules throughout the whole body of water. However, the entry of oxygen molecules through the air-water interface is much easier than the diffusion of the molecules downward through the liquid. As a result, at any time, the uppermost water layers become rapidly saturated and the rate of entry soon becomes very small. The deeper water layers also soon become starved for oxygen molecules. The result is, at any depth and at any time, the concentration gradient or the driving force is very small. The diffusion of truly stagnant water is therefore an extremely slow process that requires days or weeks to approach dissolved oxygen saturation. The process is slow because of the blocking action of molecular diffusion. Agitation of the water by stirring, by breaking the water into drops or thin layers, or by blowing bubbles of air through it will speed up this process.

2.4. Turbulent Mixing

Suppose now the water is being mixed by stirring. As a result, the stationary water layers do not exist. Let us consider a volume element of water instead. The element may have moved up to the water surface from below, remains there for a definite period of time, however small, and absorbs a relatively large amount of dissolved oxygen. The volume element then moves downward to a deeper location where it encounters a second volume element that contains very little dissolved oxygen, assuming it has never been at the water surface. For the moment at the interface between them, there exists a large concentration difference. The driving force for the molecular diffusion is therefore large and the transfer of dissolved oxygen from the first volume element to the second is very rapid. Extending this example to all of the volume elements, one can see how mixing can greatly speed up the gas transfer process. The water surface is constantly replaced by volume elements from below and hence, the blocking action of molecular diffusion is no longer present. Because of turbulent mixing and rapid diffusion, the resultant average concentration is the same at any location, within the main body of the water. So the concentration gradient or the driving force of molecular diffusion between any two volume elements, one being from the water surface and the other being remaining body of water, remains relatively high.

Note that because of turbulent mixing, the concentration gradient exists in every direction, as opposed to a preferred downward direction in the case of completely quiescent water. Therefore, with mixing, the water can become saturated with dissolved oxygen in minutes, instead of days or longer. The faster the water is mixed and the water surface renewed, the more rapid the gas transfer will be.

The rate of mass transfer under turbulent mixing can be expressed in a form similar to Fick's law,

$$n^{\mathrm{t}} = -D_{\mathrm{m}}^{\mathrm{t}} \frac{\mathrm{d}\langle C \rangle}{\mathrm{d}y} \tag{5}$$

$D_{\mathrm{m}}^{\mathrm{t}}$ is the overall mass transfer coefficient, which accounts for the combined effect of molecular diffusion and turbulent mixing (or turbulent diffusion). $\langle C \rangle$ is the time averaged concentration, and n^{t} is the overall (molecular plus turbulent) rate of mass transfer .

2.5. Air-Water Interface

Water exposed to air tends to approach, as the ultimate state, an equilibrium or saturation state. The rate of approach to equilibrium state and the actual equilibrium state are not totally independent. Under comparable conditions, the rate of approach is greater the further the system is from equilibrium.

One of the factors affecting this rate of approach to equilibrium is the formation of films at the air-water interface. These films present resistance to gas transfer. It is reported that, when the air-water interface is formed, molecules at the interface become oriented into a definite pattern, thus losing the free and random movement characteristics of the molecules in the main body of the gas and liquid. These films of oriented molecules are considered an important barrier to gas transfer.

The thickness of the film, a fictitious length, is a useful means of visualizing resistance to gas transfer. The thickness of a quiescent water film, estimated from absorption data, is of the order of 0.2 cm, whereas that of a falling water drop is about 0.0003 cm. The resistance ratio, a more useful parameter than the film thickness, is about 700 to 1 (7). The existence of the films and the rates of gas transfer through them are the subject of many theoretical studies presented in the following.

3. THEORIES OF GAS TRANSFER

3.1. Mass Transfer Equation

Gas transfer to or from a liquid is a process of mass transfer in the presence of a concentration gradient. This process may be described by the general mass conservation equation for incompressible fluids, given in vector notation as:

$$\frac{\partial c}{\partial t} + V \cdot \nabla C = -\nabla \cdot N + R \tag{6}$$

Where C and N are the concentration and mass flux vectors of the diffusing substance at time t, respectively. V is the velocity vector of the fluid, R is a source-sink term, and $V = i(\partial/\partial x) + j(\partial/\partial y) + k(\partial/\partial z)$, is the three-dimensional gradient vector.

If the gas behaves in accordance with Fick's Law,

$$N = -D_{\mathrm{m}} \nabla C \tag{7}$$

Where D_{m} is the coefficient of molecular diffusion. Equation (6) then becomes (for a constant D_{m})

$$\frac{\partial C}{\partial t} + V \cdot \nabla C = D_{\mathrm{m}} \nabla^2 C + R \tag{8}$$

If the system is stationary, i.e., $V = 0$, Equation (8) reduces to

$$\frac{\partial C}{\partial t} = D_{\mathrm{m}} \nabla^2 C + R \tag{9}$$

For steady-state mass transfer and with no chemical reaction or biological activity, i.e., without the source-sink term, the equation further reduces to:

$$\nabla^2 C = 0 \tag{10}$$

The case of mass transfer accompanied with a reaction at an interface is a special class of heterogeneous reaction problem, and will not be considered in this chapter.

For mass transfer with turbulent mixing or flow, the concentration C and the velocity vector V in the equations are replaced, respectively, by the average concentration $\langle C \rangle$ and the average velocity vector $\langle V \rangle$, and the coefficient of molecular diffusion D_{m} by the overall mass transfer coefficient $D_{\mathrm{m}}^{\mathrm{t}}$.

Consider a one-dimensional gas transfer problem, i.e., an infinite horizontal fluid of uniform thickness L_{f} has concentrations C_{A} and C_{B} at its upper and lower faces, respectively. The steady-state concentration profile in the fluid is obtained by integrating twice the equation $d^2C/dy^2 = 0$ [Equation (10)], where y is the direction into the fluid. The solution [Equation (11)] is indicating that the concentration profile in the fluid is a straight line. The mass flux is given by Equation (12).

$$C = -\frac{(C_{\mathrm{A}} - C_{\mathrm{B}})}{L_{\mathrm{f}}} y + C_{\mathrm{A}} \tag{11}$$

$$n = -D_{\mathrm{m}} \frac{\mathrm{d}C}{\mathrm{d}y} = \frac{D_{\mathrm{m}}}{L_{\mathrm{f}}} (C_{\mathrm{A}} - C_{\mathrm{B}}) \tag{12}$$

The flux, $\cap$, at either surface is given by this equation. Note that the flux is independent of position y, i.e., constant across the fluid. The thickness, L_{f}, affects inversely to the amount of gas transferred. The two-film theory discussed in the following is a practical application of Equation (12).

3.2. *Two-Film Theory*

Lewis and Whitman (8, 9) in 1924 postulated that laminar films exist at the gas-liquid interface, one liquid and one gas (Figure 5.2). These films are present regardless of the turbulent conditions in the main body of gas or liquid. Turbulence only serves to reduce the thickness of the films. If all the gas diffusing through the gas (or liquid) film must also diffuse through the liquid (or gas) film, and if the concentration becomes uniform once outside of the film region, then the basic time rate of mass transfer is:

$$n = \frac{\mathrm{d}W}{A\mathrm{d}t} = k_1(C_{\mathrm{i}} - C_{\mathrm{L}}) = k_{\mathrm{g}}(p_{\mathrm{g}} - p_{\mathrm{i}}) \tag{13}$$

Where $\mathrm{d}W/A\mathrm{d}t$ is the mass of gas passing through a unit area in a unit time. C_{i} and C_{L} are the concentrations of gas at the interface and in the main body of the liquid, respectively. p_{g} and p_i are, respectively, the partial pressures of the gas in the main body of the gas and at the interface. k_1 and k_g are the liquid film and gas film coefficients, respectively. Equilibrium between gas and liquid interface is assumed, so p_{i} and C_{i} are related by Henry's law or $C_{\mathrm{i}} = k_{\mathrm{s}} p_{\mathrm{i}}$.

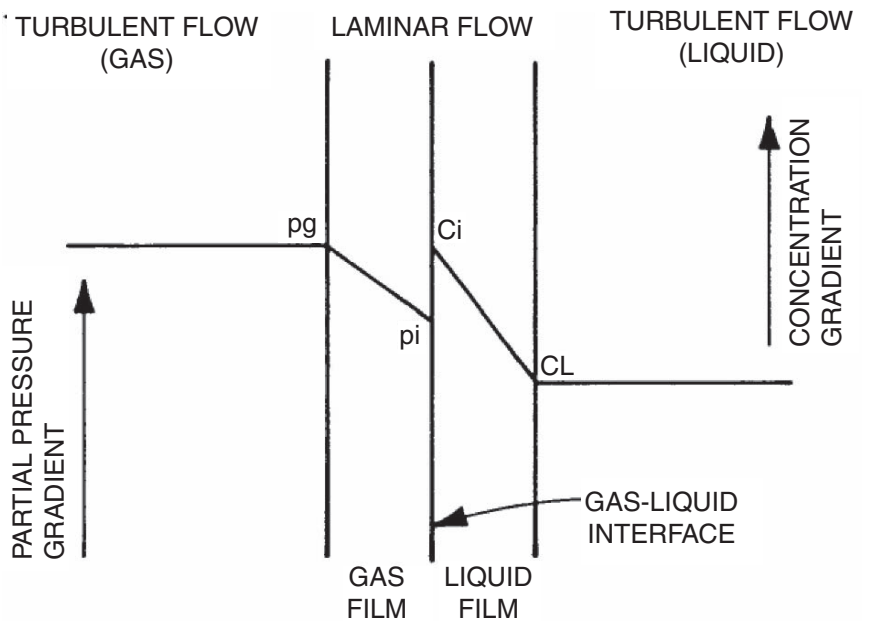

Fig. 5.2. Pressure and concentration gradients in gas and liquid films, the two-film theory.

Comparison of Equations (12) and (13) shows that the liquid film coefficient, K_L, can be expressed in terms of molecular diffusivity D_{m} and film thickness L_{f} as

$$K_{\mathrm{L}} = \frac{D_{\mathrm{m}}}{L_{\mathrm{f}}} \tag{14}$$

Equations (13) and (14) indicate that the film coefficients are dependent upon the characteristics of the liquid and the gas and the turbulent conditions, but independent of the dissolved gas in the liquid (7, 10). In real systems, concentrations and partial pressures at the interface, i.e., C_{i} and p_i are physically impossible to measure, and consequently, are the individual film coefficients.

It is therefore convenient to rewrite Equation (13) by defining overall liquid-film and gas-film coefficients K_{L} and K_{G} respectively, based on measurable quantities such as bulk liquid and gas-phase concentrations; i.e.,

$$\frac{\mathrm{d}W}{A\mathrm{d}t} = K_{\mathrm{L}}(C_{\mathrm{S}} - C_{\mathrm{L}}) = K_{\mathrm{G}}(p_{\mathrm{g}} - p_{\mathrm{s}}) \tag{15}$$

Where C_{S} is the saturation concentration and p_{s} is the corresponding partial pressure. The values of K_{L} and K_{G} are usually obtained from experimentation. K_{L} is generally referred to as the "liquid film coefficient" (see Equations (38) to (40) for relationship between overall and individual film coefficients).

The assumption of the existence of the film, and particularly the assumption of the steady state rate of transfer have been the subject of many objections which lead to the many theories discussed below.

3.3. *Penetration Model*

The penetration model was proposed by Higbie (11) and later modified by Danckwerts (12). They both consider that there exists a laminar film at the surface of the liquid with a film thickness larger than that which can be penetrated by molecular diffusion during the existence of the film. The film is constantly replaced by fresh liquid below; however, the film replacement process differs according to the two authors.

The one-dimensional unsteady-state mass transfer in a laminar layer (with no reaction), defined by Equation (9), is rewritten as:

$$\frac{\partial C}{\partial t} = D_{\mathrm{m}} \frac{\partial^2 C}{\partial y^2} \tag{16}$$

For a film of infinite depth and under the following boundary conditions,

$$\begin{aligned} C &= C_{\mathrm{L}} \quad \text{at} \quad t = 0,\ y > 0 \\ C &= C_{\mathrm{S}} \quad \text{at} \quad t > 0,\ y = 0 \\ C &= C_{\mathrm{L}} \quad \text{at} \quad t > 0,\ y = \infty \end{aligned} \tag{17}$$

The solution to Equation (16) in terms of the commentary error function is given as:

$$\mathrm{C} = C_{\mathrm{L}} + (C_{\mathrm{S}} - C_{\mathrm{L}})\ \mathrm{erfc} \left\{ \frac{\mathrm{y}}{2(t D_{\mathrm{m}})^{1/2}} \right\} \tag{18}$$

Where C is the concentration at a distance y below the surface, C_{L} and C_{S} are the concentration in the body of the liquid and the saturation concentration, respectively. t is the time elapsed since the formation of the film.

The instantaneous rate of gas transfer across the surface is

$$n = -D_{\mathrm{m}} \frac{\partial C}{\partial y}_{|y=0} = (C_{\mathrm{S}} - C_{\mathrm{L}}) \left\{ \frac{D_{\mathrm{m}}}{\pi t'} \right\}^{1/2} \tag{19}$$

and the total amount of gas transferred through the interface (unit area) during a time interval of t' is.

$$\int_0^{t'} n \mathrm{d}t = 2(C_{\mathrm{S}} - C_{\mathrm{L}}) \left\{ \frac{D_{\mathrm{m}}}{\pi t'} \right\}^{1/2} \tag{20}$$

The average rate of transfer $\langle n \rangle$ during the time t' is therefore:

$$\langle n \rangle = 2(C_{\mathrm{S}} - C_{\mathrm{L}}) \left\{ \frac{D_{\mathrm{m}}}{\pi t'} \right\}^{1/2} \tag{21}$$

Higbie considered all the surface elements as being of the same age (i.e., receiving the same duration of surface exposure) and being replaced on reaching at age t_{e}, the time of surface

renewal. Replacing t' with t_e in Equation (21) gives the average rate of transfer. Replacing the time of surface renewal, t_e, with the rate of renewal, r, where $r = 1/t_e$, the average rate of transfer becomes:

$$\langle n \rangle = \frac{2}{\sqrt{\pi}}(C_S - C_L)\{D_m r\}^{1/2} \tag{22}$$

and the overall liquid film coefficient is [by comparison with Equation (15)]

$$K_L = \frac{2}{\sqrt{\pi}}\{D_m r\}^{1/2} \tag{23}$$

Danckwerts assumed that the age, t, of any particular vertical element is a random variable having exponential distribution with its probability density function given by:

$$f(t) = re^{-rt}, t > 0 \tag{24}$$

where r is the average rate of replacement.

The rate of gas transfer over the fractional area having ages between t and $t + dt$ is given by

$$dn = -D_m \frac{\partial C}{\partial y}_{|y=0} f(t)dt = -D_m re^{-rt} \frac{\partial C}{\partial y}_{|y=0} dt \tag{25}$$

and the total rate of transfer across any unit area is

$$n = \int dn = -D_m r \int_0^\infty e^{-rt} \frac{\partial C}{\partial y}_{|y=0} dt = \sqrt{D_m r}(C_S - C_L) \tag{26}$$

and

$$K_L = \sqrt{D_m r} \tag{27}$$

Higbie's K_L is found to be 1.13 times larger than that of Danckwerts, the difference being caused by the distribution of the surface ages. Both the above models assume an infinite liquid depth. However, it is not necessary to make this assumption. Applying the Danckwerts' model to a finite depth of liquid, H, Dobbins (13) obtains:

$$K_L = \sqrt{D_m r} \tanh\left\{\frac{rH^2}{D_m}\right\}^{1/2} \tag{28}$$

In most practical cases, the tanh term is very close to unity and the assumption of infinite depth is insignificant in relation to the K_L values.

3.4. Film-Penetration Model

Dobbins (13) considered Whitman's liquid film being continuously replaced by liquid from layers beneath the surface. He proposed a model in which he applied Danckwerts' model to a

liquid film of finite thickness L_f. The boundary conditions are:

$$\begin{aligned} C &= C_L \quad \text{at} \quad t = 0, 0 < y < L_f \\ C &= C_S \quad \text{at} \quad t > 0, y = 0 \\ C &= C_L \quad \text{at} \quad t > 0, y = L_f \end{aligned} \tag{29}$$

[Compare with the boundary conditions in Equation (17).]

The solution of the model equations leads to a mass transfer coefficient given by:

$$K_L = \sqrt{D_m r} \coth \left\{ \frac{r L_f^2}{D_m} \right\}^{1/2} \tag{30}$$

When the value of the surface renewal rate, r, approaches zero, the value of K_L approaches D_m/L_f as it should for the steady-state condition.

When the value of $\{rL_f^2/D_m\}^{1/2}$ becomes greater than about 3.0, K_L is essentially equal to $\sqrt{D_m r}$. Thus, the Whitman-Lewis and the Darckwelts equations may be viewed as being special cases of Equation (30). The film-penetration model idealizes the diffusion process as taking place in two separate steps; the first step is molecular diffusion described by the film concept of Lewis and Whitman, and the second step is turbulent diffusion described by the penetration concept of Higbie and Danckwerts. The combined model has considerable merit in that it can encompass as much wider ranges of conditions than either of the film or penetration models alone. However, the model is also subject to modifications. Different renewal or mixing processes for the turbulent diffusion are considered by other investigators.

3.5. *Surface Renewal-Damped Eddy Diffusion Model*

King (14) proposed a model wherein the mass transfer near the surface is a combination effect of molecular diffusion and turbulent diffusion caused by small-scale eddies. He defined the turbulent diffusivity D' as

$$D' = ay^n \tag{31}$$

where a and n are parameters independent of time. The surface renewals are the result of large-scale eddies and the surface ages are assumed constant as Higbie's. The governing differential equation is:

$$\frac{\partial C}{\partial t} = \frac{\partial}{\partial y} \left\{ (D_m + ay^n) \frac{\partial C}{\partial y} \right\} \tag{32}$$

The solution of Equation (32) under the boundary conditions of Higbie gives, for large t,

$$K_L = a^{1/n} D_m^{(1-\frac{1}{n})} \left(\frac{n}{\pi} \right) \sin \left(\frac{\pi}{n} \right) \tag{33}$$

The model is physically more realistic than previous models. However, it involves the estimation of three parameters and it must be solved numerically in most cases.

3.6. Turbulent Diffusion Model

Kishinevsky (15) concluded that the absorption process is in most instances controlled by turbulent diffusion. His gas transfer coefficient, which considered the combined effect of molecular diffusion and turbulence, is written as:

$$K_{\mathrm{L}} = \frac{2}{\sqrt{\pi}} \left\{ (D_{\mathrm{m}} + D')r \right\}^{1/2} \tag{34}$$

where D' is the turbulent diffusivity. For high turbulence level, the turbulent diffusivity is much larger than the molecular diffusivity so that

$$K_{\mathrm{L}} = \frac{2}{\sqrt{\pi}} \left\{ D'r \right\}^{1/2} \tag{35}$$

Kishinevsky and Serebryansky (16) reported on experiments in which the measured gas transfer coefficients for hydrogen, oxygen, and nitrogen in a rapidly mixing tank are all roughly the same, as shown by Equation (35).

3.7. Other Models

Krenkel and Orlob (17) presented a model based on kinetic theory. Their model considered the frequencies and velocities of the solute molecules hitting the interface from both the liquid and the gas phases. The resulting equation expresses K_{L} as a function of $M(\mu_{\mathrm{o}}')^2/2$, where M is the molecular weight of the gas, and μ_{o}' is the normal velocity that a solute molecule must attain to be able to leave the liquid phase. Tsivoglou (18) also presented another kinetic model. His model is based on the rate difference between the entry and the loss from the interface. The gas transfer coefficient is:

$$K_{\mathrm{L}} = bL_f r \tag{36}$$

where b is the proportion of available molecules in the surface layer that actually escape, L_{f} is the thickness of surface layer from which gas molecules can escape to the atmosphere, and r is the rate of surface renewal. Fortescue and Pearson (19) considered the turbulent surface as a film of eddy cells rather than a stagnant film. The surface age is defined by a set of velocity pattern and a steady-state gas transfer condition. The resulting gas transfer coefficient, obtained by numerical solution, is

$$K_{\mathrm{L}} = 1.46 \left\{ \frac{D_{\mathrm{m}} U'}{\Lambda} \right\}^{1/2} \tag{37}$$

where U' is the longitudinal turbulent velocity, and Λ is the turbulent scale parameter.

3.8. Comparison of Gas Transfer Coefficients

The theoretical gas transfer coefficients are generally defined as a function of molecular diffusivity and turbulence. Table 5.2 lists the various gas transfer coefficient equations. The dependence on D_{m}, the molecular diffusivity, is seen to vary from an exponent of 1 for the two-film model to 0 for the turbulent diffusion model. The film-penetration model indicates

Table 5.2
List of equations for liquid-film coefficient K_L

1. Two-film theory:	$K_L = \frac{D_m}{L_f}$
2. Penetration model:	
Higbie's:	$K_L = \frac{2}{\sqrt{\pi}} \{D_m r\}^{1/2}$
Danckwerts':	$K_L = \sqrt{D_m r}$
3. Film-penetration model:	$K_L = \sqrt{D_m r} \coth \left\{ \frac{r L_f^2}{D_m} \right\}^{1/2}$
4. Surface renewal—damped eddy diffusion model:	$K_L = a^{1/n} D_m^{(1-1/n)} \left(\frac{n}{\pi}\right) \sin\left(\frac{\pi}{n}\right)$
5. Turbulent diffusion model:	$K_L = \frac{2}{\pi} \left\{ (D_m + D')\, r \right\}^{1/2}$
6. Kinetic model:	$K_L = b L_f r$

an exponent on D_m that varies from 0.5 to 1.0, whereas the surface renewal-damped eddy diffusion model shows an exponent anywhere from 0 to 1.0. It seems each model is applicable to a limited range of turbulent conditions, except the surface renewal-damped eddy diffusion model which is applicable over the entire range of turbulent conditions. Turbulence is represented by the film thickness, L_f, or by the rate of water surface renewal, r, in the various models. These parameters are no more susceptible to direct determination or measurement than the gas transfer coefficient itself. Therefore, many investigators have attempted to define the gas transfer coefficient in terms of measurable parameters. In stream reaeration, for example, its formulation is to directly relate the stream reaeration to stream velocity and stream depth. Further discussions are given in Section, Natural Reareation.

3.9. Gas-Liquid Relation

According to the two-film theory, the mass transfer in a laminar film is by molecular diffusion alone, a process much slower than the turbulent diffusion. Hence, the resistance to mass transfer is concentrated in the two films. It is also shown that the coefficients are dependent on the properties of the liquid and the gas. By substituting equivalent terms from Equations (13) and (15) and Henry's law into:

$$C_S - C_L = (C_i - C_L) + (C_S - C_i) \tag{38}$$

yields

$$\frac{1}{K_L} = \frac{1}{k_l} + \frac{k_s}{k_g} \tag{39}$$

Similarly,

$$\frac{1}{K_G} = \frac{1}{k_g} + \frac{1}{k_s k_l} \tag{40}$$

The equations show, for gas-liquid that behaves in accordance with Henry's law, the overall liquid and gas film coefficients to be independent of the gas concentration in the liquid.

A similar result has been indicated by Equation (14). Equations (39) and (40) also show the overall resistance $1/K_L$ or $1/K_G$ as the sum of the individual film resistances.

For slightly soluble gases such as oxygen and carbon dioxide, the absorption coefficient k_s is small and thus

$$\frac{1}{K_L} = \frac{1}{k_1} \tag{41}$$

This equation indicates that the major resistance occurs in the liquid film. According to Equation (14), the resistance, $1/k_1$, is directly proportional to the film thickness. Hence, reducing the thickness of the liquid film by stirring or agitation of the liquid will promote the gas transfer. For highly soluble gases such as NH_3, k_s is large, then,

$$\frac{1}{K_G} = \frac{1}{k_g} \tag{42}$$

and the major resistance occurs in the gas film. Similarly, reducing the thickness of gas film by moving or stirring the gas will promote the gas transfer.

For moderately soluble gases such as H_2S, the effect of both films are important and require both gas and liquid to be stirred or agitated to reduce the film thickness if gas transfer is to be promoted.

4. AERATION EQUATION

4.1. Significance of the Aeration Equation

Aside from the difference in the liquid film coefficient equations developed by the various models, the rate of gas transfer equation or the aeration equation has the general form [see Equations (13) and (15)]:

$$\frac{dW}{V\,dt} = \frac{dC}{dt} = K_L \left(\frac{A}{V}\right)(C_S - C) \tag{43}$$

or

$$\frac{dC}{dt} = K_L a(C_S - C) \tag{44}$$

where C is the concentration of the gas in the liquid at any time t, C_S is the saturation concentration of the gas at the interface, A is the gas-liquid interfacial area, V is the volume of the liquid, and K_L is the gas transfer coefficient, and $K_L a$ is an overall gas transfer coefficient. Because it is difficult to measure the area of gas liquid interface, A, Equation (44) is commonly employed. The following properties are indicated:

1. The rate of gas transfer is directly proportional to the concentration gradient, as Adeney and Becker reported (20). As time elapses and the gas transfer proceeds, the value of $(C_S - C)$ changes and so does the rate of gas transfer. The value of $(C_S - C)$ is thus the driving force of the gas transfer process.
2. The rate is directly proportional to the area-volume ratio, A/V. Thus, for a given volume, the rate of gas transfer is increased by increasing the area exposed, e.g., water droplets have the largest ratio.

3. The rate is also directly proportional to the overall liquid-film coefficient, which in turn depends upon the diffusivity of the gas and the film thickness or film resistance. Agitation reduces the film thickness and so increases the rate of gas transfer.
4. Temperature and pressure are important factors as they influence gas solubility and therefore the saturation concentration of gas in the liquid, C_S. Temperature also influences diffusivity and film resistance and hence the overall liquid film coefficient K_L.

Equation (44) can be integrated to yield:

$$K_L a = \frac{\ln(C_S - C_1) - \ln(C_S - C_2)}{t_2 - t_1} \tag{45}$$

where C_1 and C_2 are the dissolved gas concentrations of the liquid under aeration after elapsed times t_1 and t_2, respectively.

The overall gas transfer coefficient, $K_L a$, is the product of K_L and A/V, and thus both the values of $K_L a$ and K_L are equally affected by the temperature of the liquid, the characterization of contaminants and the intensity of mixing, as discussed in the following.

4.2. *Influencing Factors*

1. Temperature: The effect of temperature change on the rate of gas transfer is two-fold: (1) increasing the temperature makes the gas less soluble (lower C_S), and thus lowers the rate of gas absorption but increases the rate of gas release, and (2) increasing the temperature reduces the film resistance but increases the molecular diffusivity, thus increasing the gas transfer coefficient, which makes both absorption and release of gas more rapid. The temperature effect is usually expressed as

$$K_{L(T)} = K_{L(20)}\theta^{(T-20)} \tag{46}$$

where $K_{L(T)}$ and $K_{L(20)}$ are respectively the coefficients at temperature T°C and 20°C. and θ is the temperature coefficient. The value of θ has been reported, from experimental data, to vary from 1.016 to 1.047 with a common value of 1.024 generally used (21–23).

Tsivoglou (18) showed based on kinetic theory, the temperature relation:

$$\frac{K_{L(T_1)}}{K_{L(T_2)}} = \frac{C_S(T_2)}{C_S(T_1)} = \theta^{(T_1 - T_2)} \tag{47}$$

where T_1 and T_2 are two different water temperatures, $K_{L(T1)}$ and $K_{L(T2)}$ are the respective values of K_L, and $C_S(T_1.)$ and $C_S(T_2)$ are the respective values of the saturation concentration. The values of θ calculated by Tsivoglou from known dissolved oxygen saturation concentration in freshwater for temperatures from 0 to 30°C range from 1.018 to 1.026 with a mean value of 1.022. Exactly the same value was also found for water containing large amounts of chlorides.

2. Turbulence: Turbulence is generated from the motion of liquid. For example, rising bubbles induce velocity gradients in a liquid to produce mixing and surface turbulence, whereas the turbulence within the falling droplets is caused by the continuous motion originating in the droplet produced from a device such as a spray nozzle. Turbulent mixing: (1) minimizes concentration differences within the main body of the water and thus maintains higher concentration gradient across the gas-liquid interface, and (2) increases the rate of surface renewal and produces a larger area-volume ratio. As a result, higher gas transfer efficiency is achieved. Turbulence also reduces film thickness or the resistance to molecular transfer. The film thickness computed from gas diffusivity data showed a relation with this effect (24).

3. Waste Constituents: Waste constituents in water, similar to the turbulence, also affect the liquid-film coefficient and the area-volume ratio. The influence to the coefficient is on the diffusivity in this instance. Waste constituents such as surface active agents or surfactants (soaps, synthetic detergents, organic acids, and so on) will concentrate at the liquid-air interface and, as a result, the film of adsorbed surfactant molecules at the interface provides a further barrier to molecular diffusion (23, 25). Its effect is a function of the waste concentration. When present in minute amounts, considerable reductions of the rate of gas transfer to water are reported. Kehr (26) reported that 6 mg/L of soap in water in an aeration test channel, reduced the reaeration rate by about 50%. At high concentrations, the effect is reduced and K_L tends to increase slightly. The effect on the area-volume ratio is particularly significant in bubble aeration. Bubble sizes are seen to decrease with increasing concentrations of surfactants, which influence the formation of bubbles (and droplets) by decreasing the surface tension in water. The increase of the K_L value at high surfactant concentration has been reported primarily to result from the increased area-volume ratio (27, 70).

   In waste treatment application the change in transfer rate in the presence of surfactants or wastes is defined by the waste correction factor α, as

$$\alpha = \frac{K_L a \text{ of wastewater}}{K_L a \text{ of cleanwater}} \tag{48}$$

   The values of α for various aeration systems and wastewaters have been reported to vary widely from about 0.5 to over 1.3. with values around 0.85 to 0.90 being common for domestic sewages (23, 28).

   The solubility of oxygen in water is also affected by the impurities, such as salts, or other dissolved solids, present in the water. A constant, β, is defined as follow:

$$\beta = \frac{C_S \text{ of wastewater}}{C_S \text{ of cleanwater}} \tag{49}$$

   The value of β generally ranges from 0.9 to 1.0 for sewage.
4. Gas Partial Pressure: Increasing the partial pressure of the gas in the air will produce in proportion an increase of the value of saturation concentration, as expressed by a simple relation:

$$C_{S1} = C_{S2}\frac{p_1}{p_2} \tag{50}$$

   where C_{S1} and C_{S2} are, respectively, the saturation concentrations at partial pressures p_1 and p_2. The increase in the gas partial pressure is beneficial for gas absorption, but disadvantageous for gas release, or vice versa.

   To increase the partial pressure, compressed air or pure gas can be added to the air supplied to an enclosed aerator. To reduce the partial pressure, vacuum or inert gas can be applied or added, respectively. For gas release (the later case), adequate ventilation of the enclosed structure is essential. Without ventilation, the partial pressure would increase as the removal continued thus slowing the rate of release. Adequate ventilation is also required for gas absorption. It prevents the lowering of the gas partial pressure and also adds the advantage of air turbulence to the system. Forced-draft aerators are designed for such a purpose.

4.3. Natural Reaeration

Many of the theories concerning the mass transfer of a gas to and/or from a liquid given in Section 3, were developed from the analysis of stream reaeration. In a surface water body,

when an oxygen sink such as the process of biochemical oxidation of organic materials exists, the water body would become undersaturated with oxygen. To replenish oxygen to the water, the process of natural reaeration takes place. Oxygen is transferred from the limitless resources of atmosphere into the water at the water surface, i.e., the air-water interface. This interface is constantly being replaced by the water below, because of the turbulent mixing of the (flowing) water, and the higher the rate of this surface renewal the greater the ability of the water body to absorb oxygen from the atmosphere. However, in any stream the rate of oxygen replenishment or the stream reaeration capacity is finite because of the stream's physical characteristics, which control the degree of turbulent mixing in the stream. This, in effect, limits a stream's ability to receive and assimilate oxygen-depleting wastes and in turn dictates the necessary degree of waste treatment required and the associated treatment costs. Hence, the natural reaeration capacity of a surface water body is a valuable resource, and is a vitally needed design parameter if the waste treatment facility is to be both adequate and economical.

4.3.1. Stream Reaeration Coefficient

For turbulent water system in which oxygen saturation deficits, D, exist but without any measurable vertical concentration gradients (29). One can write

$$D = (C_S - C) \tag{51}$$

where C_S is the dissolved oxygen saturation, and C is the average dissolved oxygen. The basic aeration equation, Equation (44) is rewritten as:

$$\frac{dD}{dt} = -K_2 D \tag{52}$$

where K_2 is the reaeration coefficient, or the rate constant for the absorption of oxygen from the atmosphere. Equation (52) states that the rate absorption is directly proportional to the saturation deficit. i.e., the higher the saturation deficit the greater the rate of reaeration (20, 30). The solution of Equation (52) is :

$$D = D_o e^{-K_2 t} \tag{53}$$

where D_o is the initial dissolved oxygen deficit at $t = 0$. Integrating Equation (52) between times t_1 and t_2 and solving for K_2, yields:

$$K_2 = \frac{\ln D_1 - \ln D_2}{t_2 - t_1} \tag{54}$$

Equation (54) can be used to obtain the reaeration coefficient for waters under reaeration process only. For polluted river water, the oxygen balance method of Streeter and Phelps (30) is used to evaluate the stream reaeration coefficient.

The Streeter and Phelps' differential equation describing the combination of deoxygenation and reaeration is:

$$\frac{dD}{dt} = K_1 L - K_2 D \tag{55}$$

where L is the concentration of organic matter in BOD (biochemical oxygen demand), and K_1 is the deoxygenation coefficient. The integrated solution of Equation (55) is called oxygen sag equation and is used in the analysis of the dissolved oxygen balance in a polluted stream. By substituting measured or known BOD and DO (or D) values and the K_1 value, one can solve for the reaeration coefficient from Equation (55) or from the oxygen sag equation.

Because there are errors involved in the estimations of the known values processes other than biochemical oxidation and reaeration that are not considered by Equation (55), the reaeration coefficient computed may not be considered accurate. Other authors have expanded the fundamental Streeter and Phelps equation to include other sources and sinks of DO in the oxygen balance, modified the assumption of steady-state condition to unsteady-state, and so on to increase the accuracy of estimation (31, 32).

4.3.2. Direct Field Measurement of the Reaeration Coefficient

The effects of the various oxygen sources and sinks on the determination of stream aeration coefficient by oxygen balance technique, can be eliminated by using the method of gaseous tracer measurement in the stream. The gas transfer rate of the inert, radioactive krypton-85 gaseous tracer from water to atmosphere is independent of the complex DO balance of the stream, and it is correlated to the atmospheric oxygen transfer rate of the stream (18, 33, 34).

Consider a dissolved tracer gas, krypton-85, which has been added to the water. The amount of krypton-85 present in the atmosphere above the water can be taken to be zero, and so the value of krypton-85 saturation concentration in water, the transfer is from the water to the atmosphere or the tracer gas is steadily lost from the water. The desorption of the tracer gas can be described by Equation (53), rewritten as:

$$C_{\mathrm{kr}} = C_{\mathrm{kr_o}}\, e^{-(K_2)_{\mathrm{kr}} t} \tag{56}$$

where C_{kr} is the concentration of the dissolved tracer gas remaining in the water at time t, $C_{\mathrm{kr_o}}$ is the concentration at $t = 0$, and $(K_2)_{\mathrm{kr}}$ is the gas transfer coefficient for the tracer gas.

It can be shown that under identical hydraulic conditions, the following relations exist:

$$\frac{(K_2)_{\mathrm{A}}}{(K_2)_{\mathrm{B}}} = \frac{(D_{\mathrm{m}})_{\mathrm{A}}}{(D_{\mathrm{m}})_{\mathrm{B}}} = \frac{(d)_{\mathrm{B}}}{(d)_{\mathrm{A}}} \tag{57}$$

where $(K_2)_{\mathrm{A}}$ and $(K_2)_{\mathrm{B}}$ are the values of K_2 for the two different gases A and B. D_{m} is the coefficient of molecular diffusivity, and d is the molecular diameter. The above relations have also been demonstrated in an extended series of experiments involving a number of different gases (hydrogen, helium, nitrogen, carbon dioxide, oxygen, radon, and krypton).

Equation (57) states that, under identical conditions of turbulent mixing, the ratio of gas transfer coefficients for two different gases is just equal to the ratio of their molecular diffusivities or to the inverse ratio of their molecular sizes. The ratio for the krypton-85 gas and the oxygen gas is found, both experimentally and theoretically, as

$$\frac{(K_2)_{\mathrm{kr}}}{(K_2)_{\mathrm{ox}}} = 0.83 \tag{58}$$

where $(K_2)_{ox}$ is the oxygen transfer coefficient or the reaeration coefficient (in stream). It is also noted that the numerical constant, 0.83, given in Equation (58) has been demonstrated to be independent of the degree of turbulent mixing, independent of the directions in which the two gases happen to be moving, and independent of temperature within the range 10 to 30°C.

In field tracer application, samples collected from points, one upstream and one downstream, are analyzed for krypton-85 and the values at the moment of maximum tracer concentration at each location are used in Equation (54) to compute $(K_2)_{kr}$, i.e.,

$$(K_2)_{kr} = \frac{\ln(C_{kr})_1 - \ln(C_{kr})_2}{t} \tag{59}$$

where $(C_{kr})_1$. and $(C_{kr})_2$ are the maximum dissolved krypton-85 concentrations at the two sampling points, and t is the time-of-travel between the two sampling points. The oxygen transfer coefficient $(K_2)_{ox}$ (or the reaeration coefficient K_2) for that stream reach can then be obtained directly from Equation (58).

In practice, the field application involves the simultaneous releases of a mixture of three tracers: (1) dissolved krypton-85, (2) tritium in the form of tritiated water molecules, and (3) a fluorescent dye. The fluorescent dye provides an accurate measure of the time of flow between two sampling points, and also indicates when to sample the other two radioactive tracers. The tritium tracer that is not absorbed on the stream bed or otherwise lost in any significant amount provides an accurate measure of dispersion of the tracer dose or the correction factor for the effects of dispersion. Because the dissolved krypton-85 also undergoes exactly the same dispersion as the tritiated water, the concentration of krypton-85 measured at the downstream point, $(C_{kr})_2$, is corrected for the loss owing to dispersion or

$$(C_{kr})'_2 = (C_{kr})_2 \frac{(C_{tr})_1}{(C_{tr})_2} \tag{60}$$

where $(C_{kr})'_2$ is the concentration of krypton-85 at downstream sampling point if only the transfer from water to atmosphere takes place. Substituting $(C_{kr})'_2$ into Equation (59) in place of $(C_{kr})_2$, then

$$(K_2)_{kr} = \frac{\ln(C_{kr}/C_{tr})_1 - \ln(C_{kr}/C_{tr})_2}{t} \tag{61}$$

where $(C_{kr}/C_{tr})_{1,2}$ are the concentration ratios of krypton-85 and tritium in the samples taken at the time of the dye peaks at sampling points 1 and 2, and t is the time of flow between the two points. The conversion to $(K_2)_{ox}$ or K_2 can then be made using Equation (58).

The direct measuring method using radioactive tracer requires cautions in application in relation to permissible environmental limits and possible exposure of personnel handling the tracer dose. Lately, a modified tracer technique using nonradioactive tracers, such as ethylene or propane, was developed, and its preliminary application indicates that it is a promising alternative (35).

The direct and independent field measurements of stream reaeration coefficient may provide data that can be used to evaluate the accuracy or dependability of any of the available reaeration coefficient predictive equations (to be discussed in the following), or to select among them.

4.3.3. Reaeration Coefficient Prediction Equations

Note that the reaeration coefficient K_2 is equivalent to $K_L a$ (or $K_L A/V$) in aeration equations. As described for $K_L a$ the value of the coefficient K_2 is obviously dependent upon the degree of turbulent mixing or the rate of surface renewal of the water body. These parameters are no more susceptible to direct determination or measurement than the reaeration coefficient itself. Many investigators have therefore attempted to define the reaeration coefficient in terms of measurable stream parameters such as the velocity and depth of flow (36–43). Longitudinal dispersion, hydraulic or energy gradient are also included in some prediction equations (44–46).

Because the ratio of area to volume for natural stream is readily defined as $A/V = 1/H$ the reaction coefficient can be defined as:

$$K_2 = K_L a = K_L \frac{A}{V} = \frac{K_L}{H} \tag{62}$$

In application of Equation (62), one should note that the ratio of area to volume is properly regarded as the reciprocal of the average depth of water, H, only under conditions of homogeneous mixing. In a large slow-moving river, it is probable that the whole body of water is not homogeneously mixed, and thus the average depth of flow is not a measure of the depth that is effective in terms of surface replacement or reaeration. Therefore, if the average depth as defined by water surface area to water volume is considered in Equation (62), the water mixing conditions are properly reflected by the value of K_L. A plot of the reaeration coefficient versus the depth of flow and river flow conditions is shown in Figure 5.3. The water mixing conditions are defined by ranges of flow velocities which may be expected during low-flow periods, and by the water surface characteristics.

The general form of the predictive equation for reaeration coefficient is

$$K_2 = \frac{cU^n}{H^m} \tag{63}$$

where c, n, and m are constants, U is the velocity of flow, and H is the depth of flow. The value of c includes the factors of the roughness of the bottom, the slope of the stream, and the number of changes in flow direction, etc. Some of the typical predictive equations are presented below.

Streeter and Phelps (30) developed an empirical formula that relates the reaeration coefficient to the physical characteristics of the stream,

$$K_2 = \frac{2.3\, cU^n}{H'^2} \tag{64}$$

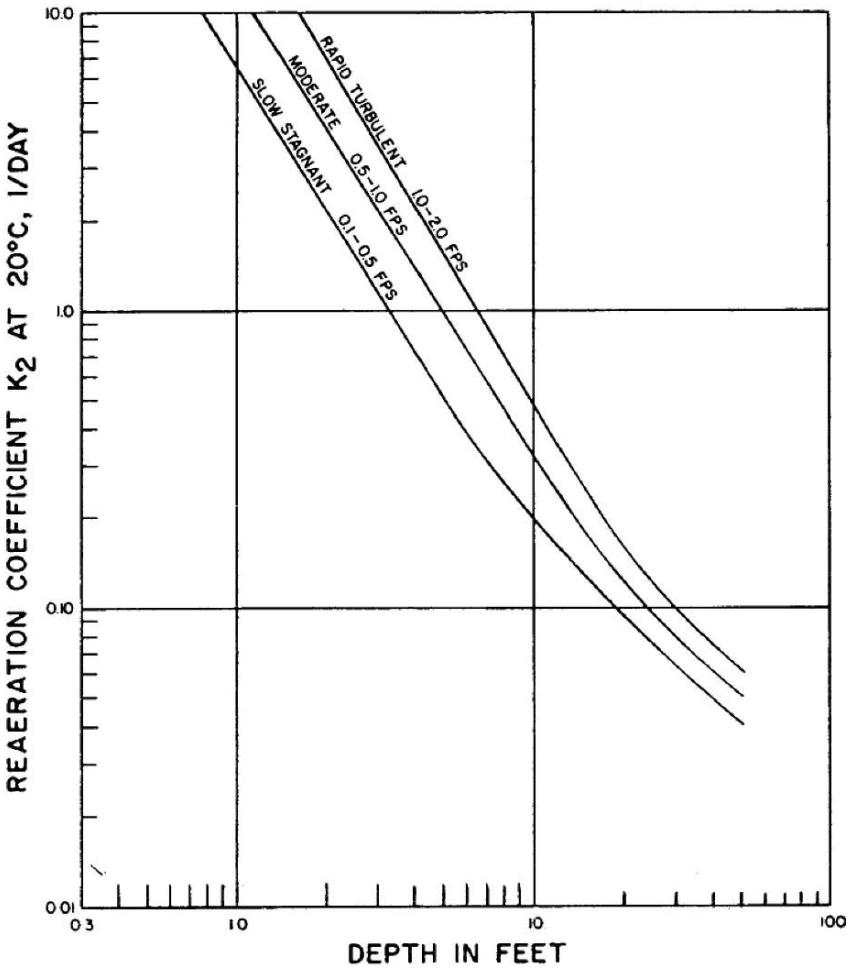

Fig. 5.3. Reaeration coefficient (K_2) as a function of depth.

where H' is the depth above minimum low water stage. The constant c relates to the irregularity of the river channel and the slope of water surface. The values of c for the Ohio River range from 0.2 to 131. The values of n range from 0.57 to 5.40.

O'Connor and Dobbins (37), by theoretical derivation, developed an equation for K_2 applicable to most of the natural streams as follows:

$$K_2 = \frac{{D_L}^{1/2} U^{1/2}}{H^{3/4}} \tag{65}$$

where D_L is the coefficient of oxygen molecular diffusion in water (0.81×10^{-4} ft^2/h at 20°C), U is the average stream velocity and H is the average depth. If U is in feet/second

and H in feet, then

$$K_2(1/d) = \frac{12.9U^{1/2}}{H^{3/4}} \tag{66}$$

The average depth can be taken as the ratio of the volume of water in a reach to the surface area or the ratio of cross-sectional area to width.

Churchill et al. (39) at, the Tennessee Valley Authority, conducted an extensive river survey under excellent experimental conditions. The river water was initially low in dissolved oxygen and free from organic pollution, and the stream flow control was possible. The equation developed for K_2 is:

$$K_2 = \frac{11.56U^{0.969}}{H^{1.673}} \tag{67}$$

where

K_2 = reaeration coefficient, day^{-1},
U = velocity of flow, ft/s
H = average depth, ft.

Several similar types of predicting equations were obtained by researchers using their data and the available reported data by others:

Owens, Edwards, and Gibbs (40):

$$K_2 = \frac{21.62U^{0.67}}{H_1{}^{0.85}} \tag{68}$$

Langbein and Durum (41):

$$K_2 = \frac{7.59U}{H^{1.33}} \tag{69}$$

Isaacs and Gaudy (42):

$$K_2 = \frac{8.6U}{H^{1.5}} \tag{70}$$

Same units as Churchill's are used.

5. SURFACE AERATION

5.1. Introduction

Mechanical surface aerators transfer atmospheric oxygen to waters by generating a high degree of water surface turbulence and a high rate of contact surface renewal. Water is lifted above the water surface by the pumping action of the aerator blades and discharged into the atmosphere in dispersed droplets with thin films. When the water spray splash onto the water surface, it causes turbulence at the water surface and also entrains air bubbles into water body. As a result of these actions, large air-water interfaces is attained, the oxygen diffusion takes place. The agitation and pumping actions of the aerator provide the water content with

turbulent mixing, water circulation, and continuous surface renewal. These high degrees of air-water contact, content mixing, and surface renewal result in an efficient oxygen transfer from the surface atmosphere to the water body.

In both water and wastewater treatment applications, surface aeration can be employed to promote transfer of gases or volatile compounds between the gas phase and the liquid phase. However, surface aeration has its most important application in the aerobic biological wastewater treatment. In the processes of biochemical oxidation and stabilization of organic materials, dissolved oxygen is consumed by microorganisms. The principal functions of aeration are to supply oxygen to wastewater for microbial use and to provide turbulent mixing to the entire liquid contents. The mixing keeps the biological flows in uniform suspension and intimate contact with the incoming food supply.

5.2. Types of Surface Aerators

Among various types of aeration equipment for wastewater treatment applications, mechanical surface aerators have demonstrated that they are efficient in oxygen transfer and turbulent mixing, and the equipment is simple and easy to install, operate, and maintain. Mechanical surface aerators of many types and designs have been widely used in activated sludge aeration tanks, aerobic sludge digesters, aerated lagoons, and oxidation ditches.

Bladed surface aerators have a surface impeller mounted on a vertical shaft. The action of the bladed surface aerators is similar to that of a low-lift high-flow-capacity pump. A large volume of water is pumped above the water surface and sprayed radially, causing a high degree of continuous air-water contact. The oxygen transfer efficiencies of bladed surface aerators are generally about 3 to 4 lb of oxygen transferred per shaft horsepower per hour under standard conditions. The operating speeds are commonly between 30 and 60 rpm. The oxygen transfer capacity of aerators can be varied to meet the system demands by either changing the speed of the multiple speed drive unit, or by adjusting the submergence of the impeller. Bladed surface aerators have been used in aeration tanks with a depth of up to 15 ft. With the use of a lower mixing impeller near the bottom of the tank, or a vertical draft tube extending from the water surface to near the bottom of the tank, the maximum water depth can be increased to 15 to 17 ft. Combination type turbine aerators have been installed in the deep aeration basins with a depth of up to 18 to 20 ft. In addition to a surface impeller, aerators are equipped with a bottom impeller and an air sparger. Compressed air is introduced to the bottom of the tank from a pipe or a sparge ring located beneath the lower impeller and the rising air is dispersed into the liquid by the shearing and mixing actions of the lower impeller.

Brush aerators, or rotor aerators, consist of a large number of blades, discs, paddles, or brushes mounted on a large diameter horizontal shaft. The rotating speeds of brush aerators are generally between 60 and 120 rpm. The rotor pushes and throws the water, causing a spray of water at the aerator and a turbulent flow leaving the aerator. Brush aerators are commonly used in the oxidation ditch, usually of a depth of 10 to 15 feet and they have also been used in the activated sludge aeration tanks installed along one side of the tank.

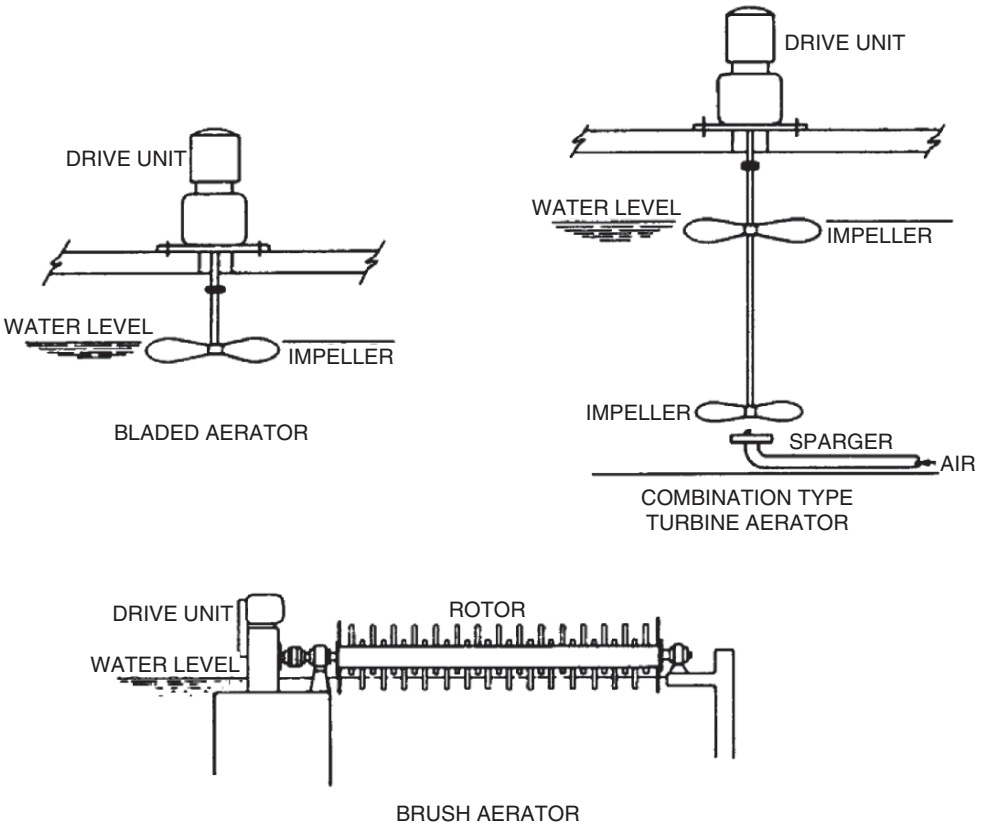

Fig. 5.4. Mechanical aerators.

Figure 5.4 shows three basic types of mechanical surface aerators. A detailed illustration of the various types and designs of commercially available mechanical surface aerators can be found in reference (47).

5.3. Techniques for Surface Aerator Performance Test

Oxygen transfer efficiency of a given aerator can be expressed by the value $K_L a$. the overall oxygen transfer coefficient. Because $K_L a$ is a function of the geometry of aeration basin, which varies from one installation to another, so a full-scale field performance test must be conducted to determine the actual oxygen transfer efficiency of an aerator installation. Various techniques have been employed to evaluate the efficiency of aeration devices (23, 28, 48–51). These include: (a) unsteady-state aeration of deoxygenated clean water, (b) steady-state aeration of activated sludge mixed liquor, (c) steady-state aeration of a continuous flow system of deoxygenated clean water (by addition of a sulfite solution), (d) unsteady-state aeration of deoxygenated activated sludge mixed liquor, (e) deaeration of krypton-85 gaseous tracer added to clean water. The first two of these methods are commonly applied in the field performance evaluation of aerator installations and are described here.

5.3.1. Unsteady-State Aeration of Deoxygenated Clean Water

Performance testing of aerators with clean water in a batch aeration basin is based on Equation (44) as:

$$\frac{dC}{dt} = K_L a(C_S - C) \tag{44}$$

The water under aeration is observed for the changes in dissolved oxygen concentration with aeration time, starting with a dissolved oxygen (DO) level close to zero to a level close to 90% of the saturation DO level, C_S, or higher. Because the values of $K_L a$ and C represent the uniform or average values for the entire liquid contents, hence a condition of complete mixing throughout the aeration tank is desirable.

Fresh water in the aeration tank is first deoxygenated by adding sodium sulfite and cobalt chloride, or cobalt sulfate, as a catalyst to the water. The amount of sodium sulfite added shall be in excess of the theoretical 7.9 mg/L sodium sulfite for each mg/L of dissolved oxygen. The progress of reaeration of the deoxygenated water is then measured and recorded. A semilogarithmic plot of dissolved oxygen deficits versus aeration times and then a line of best fit can be used with the data. The value of $K_L a$ is given as the slope of the straight line, as shown in Figure 5.5. The oxygen transfer efficiency of the aerator can now be computed by using this equation:

$$N = \frac{K_L a(C_S - C)W \times 10^{-6}}{P} \tag{71}$$

where N is the oxygen transfer efficiency of aerator in lb O_2/hp-h, C is the dissolved oxygen concentration of any desired operating level in mg/L. W is the weight of water under aeration in pounds, and P is the shaft horsepower of aerator.

The shaft horsepower of aerator can be determined by using a watt meter, or by measuring the incoming line amperage and voltage, and applying the following equations:

direct current

$$P = \frac{I \times E \times e}{746} \tag{72a}$$

single-phase, alternating current

$$P = \frac{I \times E \times PF \times e}{746} \tag{72b}$$

three-phase, alternating current

$$P = \frac{I \times E \times PF \times e \times 1.73}{746} \tag{72c}$$

where I is the amperage, E is the voltage, PF is the power factor and e is the efficiency of the motor and gear reducer.

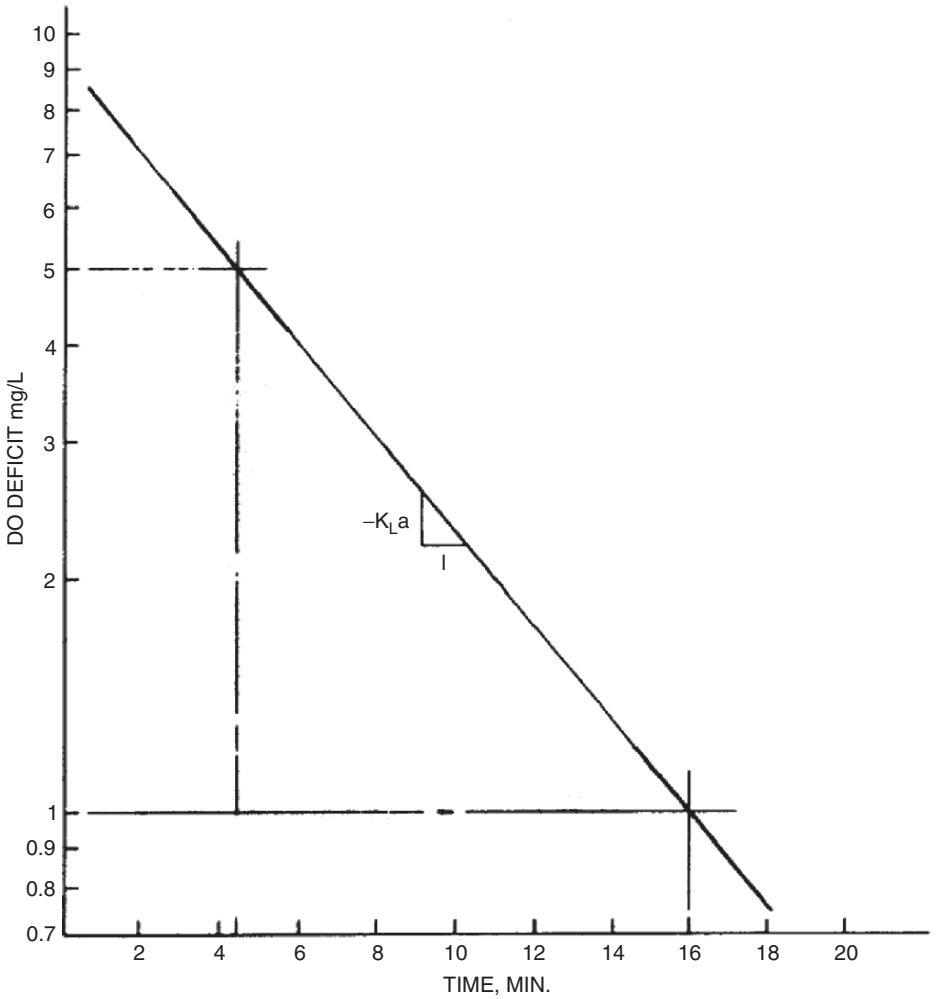

Fig. 5.5. Determination of K_La.

5.3.2. Example

A field performance test of a 30 hp surface aerator in a tank containing 163,000 gal of tap water with a water temperature of 19°C resulted in a straight line plot, as shown in Figure 5.5. Using Equation (45):

$$K_L a = \frac{\ln(5) - \ln(1)}{15.9 - 4.4} = 0.14\,\text{min}^{-1} = 8.4\,\text{h}^{-1}$$

The electrical data recorded on the three-phase, AC power supply are: average voltage = 475, and average amperage = 32.5. The drive unit at this load level has a motor efficiency of 89.5%, a power factor of 86%, and a gear reducer efficiency of 93%. The aerator shaft

horsepower can be computed by Equation (72c),

$$P = \frac{32.5 \times 475 \times 1.73 \times 0.86 \times 0.895 \times 0.93}{746} = 25.6\,\text{hp}$$

The oxygen transfer efficiency of the aerator at zero DO level ($C = 0$) is determine from Equation (71):

$$N = \frac{K_L a(C_S - C)W \times 10^{-6}}{P} \tag{71}$$

$$N = \frac{8.4 \times (9.4 - 0) \times 163{,}000 \times 8.34 \times 10^{-6}}{25.6} = 4.2\,\text{lb O}_2/\text{h/shaft hp}$$

The cobalt catalyst applied in sulfite oxidation commonly amounts to a concentration of about 2 to 5 mg/L. However, it has been found that cobalt at that level seriously interferes with the Winkler method of dissolved oxygen determination (52). With cobalt interference, the values of dissolved oxygen concentration determined by Winkler method are higher than the actual dissolved oxygen concentrations of water. Furthermore, if a theoretical value of C_S is used in the calculation for $K_L a$ a value of 10% to 50% higher than the actual value for $K_L a$ would result. It has also been demonstrated, however, that a cobalt concentration of 0.05 mg/L in aeration tank is sufficient and presents no significant interference in the Winkler determination.

In analysis of the data, it has been a common practice to consider only the data points between 10 and 70, or 20% and 90% of oxygen saturation (47). It has been shown that truncation of dissolved oxygen data up to 20% of saturation value does not affect the accuracy of $K_L a$ determination, but on the other hand, truncation of dissolved oxygen data at the higher end, does result in a decreased accuracy in $K_L a$ determination (53). The value of C_S has sensitive effect on the computed values of dissolved oxygen deficits at the higher levels of oxygen saturation, it is important that an accurate value of oxygen saturation concentration is used in the parameter estimation of oxygen transfer efficiency.

5.3.3. Steady-State Aeration of Activated Sludge Mixed Liquor

With activated sludge mixed liquor under aeration, the rate of oxygen transfer can be estimated by using the following relationship:

$$\frac{dC}{dt} = K_L a(C_{SW} - C_t) - r_o \tag{73}$$

where C_{SW} is the oxygen saturation concentration of the wastewater, r_o is the oxygen uptake rate, and C is dissolved oxygen concentration of mixed liquor.

Under steady-state conditions, the rate of oxygen input equals the rate of oxygen use, and $dC/dt = 0$, which yields:

$$K_L a = \frac{r_o}{C_{SW} - C} \tag{74}$$

The oxygen uptake rate can be measured by using various techniques (28, 47). These include: off gas analyses on the carbon dioxide produced or oxygen consumed, and polarographic or galvanometric measurement of dissolved oxygen decrease.

The measurement for oxygen uptake rate for a sample of mixed liquor must be conducted rapidly. Common methods applied in the field determination of oxygen uptake rate involve measuring the changes in dissolved oxygen concentration with time using a dissolved oxygen probe. The mixed liquor samples may be preaerated for a short period of time, if the initial dissolved oxygen concentration is low. By using unsteady-state technique, a small volume of mixed liquor is collected in a small batch tank and put under gentle agitation, but without aeration. The sample is measured for the rate of change in dissolved oxygen level by using DO probe. By using steady-state technique, a constant flow of mixed liquor sample is fed to a small test tank where it is agitated, and the DO level is determined. An additional small vessel upstream from the test tank may be included to preaerate the sample. The oxygen uptake rate can be calculated by using the following equation for the steady-state condition,

$$r_o = \frac{Q(C_1 - C_2)}{V} \tag{75}$$

where Q is the flow rate of mixed liquor sample in gallons per hour, V is the volume of mixed liquor in the test tank in gallons, and C_1 and C_2 are the DO concentrations in the influent and in the test tank, respectively.

5.3.4. Standard Conditions

The value of oxygen transfer efficiency of an aerator obtained under a set of specific test conditions is often converted to an equivalent value based on the standard conditions. A body of water at standard conditions is defined as: at 20°C temperature, zero dissolved oxygen, free of contamination, and under 1 atmosphere of pressure. The overall conversion equation is:

$$N_o = \frac{NC_{s(20°C)}}{\alpha\theta^{(T-20)}\{(\beta(p/p_o)C_S) - C\}} \tag{76}$$

where N_o and N are, respectively, the overall oxygen transfer efficiencies at standard conditions and test conditions, and p_o and p are, respectively, the oxygen partial pressures at standard conditions and test conditions). $C_{s(20°C)}$ is the saturation oxygen concentration at 20°C. Other notations have been previously defined.

5.3.5. Example

Steady-state aerator performance test gave the following data: volume of mixed liquor under aeration = 150,000 gal, water temperature = 22°C, $\alpha = 0.9$, $\beta = 0.91$, $p/p_o = 1$ and $C_{SW} = 8.01$ mg/L. The oxygen uptake rate of mixed liquor was determined to be 25 mg/L/h, and the dissolved oxygen concentration of mixed liquor was 2.60 mg/L.

$$K_La = \frac{r_o}{C_{SW} - C} \tag{74}$$

$$K_La = \frac{25}{8.01 - 2.6} = 4.26\,h^{-1}$$

The oxygen transfer rate under test conditions is

$$N = (150{,}000\,\text{gal})\frac{\text{mgal}}{10^{-6}\,\text{gal}}\left(8.01\frac{\text{mg}}{L} - 2.6\frac{\text{mg}}{L}\right)\left(8.34\frac{\text{lb}}{\text{mgal}\frac{\text{mg}}{L}}\right)\left(\frac{4.62}{h}\right) = 31.3\frac{\text{lb O}_2}{h}$$

The oxygen transfer rate under standard conditions is

$$N_\text{o} = \frac{NC_{\text{s}(20^\circ\text{C})}}{\alpha\theta^{(\text{T}-20)}\{(\beta(\text{p}/p_\text{o})C_\text{S}) - C\}} \tag{76}$$

$$N_\text{o} = \frac{31.3 \times 9.17}{0.9 \times 1.024^{(22-20)}\{(0.91(1) \times 8.8) - 2.6\}} = 56.5\,\text{lb O}_2/\text{h}$$

5.4. *Surface Aerator Design*

In sizing aerators for an aeration system, some basic factors to be considered are: the oxygen requirements of the process, the characteristics of wastewater and the anticipated field conditions, the requirements for solids suspension and mixing, the geometry of the basin, and the oxygen transfer efficiencies of aerators. Aerators selected must meet both requirements for oxygen transfer and turbulent mixing of the system. Oxygen transfer efficiencies of mechanical surface aerators in most aeration applications range from 2 to 4 lb of oxygen transferred per brake horsepower per hour at standard conditions. Oxygen transfer efficiencies of greater than 5 lb/hp/h have been reported. Equation (76) can be used to estimate the oxygen transfer capacity of an aerator for the design field conditions, and the total horsepower and the number of aerators required can be estimated for an aeration system.

In the activated sludge process and some aerated lagoons, it is essential that an adequate circulation and turbulent mixing of the entire liquid contents be provided to keep the biological flocs in uniform suspension and in intimate contact with the incoming food supply (raw wastewater). In wastewater treatment plants with activated sludge process, the power levels are determined by the requirement for oxygen transfer capacity and are generally greater than the requirement for solids mixing. In some aerated lagoons applications, the sizes of aerators are often controlled by the requirement for solids mixing.

The requirements for mixing capacity of aerator are commonly dealt with by considering: a minimum power input level of 0.5 to 1.0 hp/1000 ft^3 of aeration tank volume (54, 55), a minimum bottom scouring velocity of 0.5 to 1.0 ft/s, the required pumping capacity of aerator based on the geometry of aeration tank and the spacing of aerators, or an allowable maximum variation from the average concentration of solids for all locations in the tank.

5.5. *Artificial Instream Aeration*

The process of natural stream reaeration is quite slow for most streams. Therefore, when organic pollution loads exceed the assimilative capacity of the stream, the dissolved oxygen content in the stream is reduced to below the stream's dissolved oxygen standard and may even become depleted. Artificial instream aeration can be applied to accelerate the rate of oxygen supply to the water body; thus the levels of dissolved oxygen can be maintained above the specified minimum allowable level for the stream. Studies on the instream aeration of a

polluted river have shown that the instream aeration can be an effective and economical means of meeting the dissolved oxygen standard of the stream (29, 56, 57).

Because the oxygen transfer rate and the oxygen transfer efficiency of an aerator depend directly on oxygen deficit, it will be most economical to add the oxygen when the dissolved oxygen (DO) level has fallen to the lowest acceptable level. In addition, it will only be economical to increase the DO level to some value less than saturation at a given location. When the DO level again falls to the lowest acceptable level because of the higher oxygen consumption of pollutants than the oxygen replenishment by natural reaeration, a second aerator may be installed to maintain a high enough DO level. The mathematical model simulation technique is a very useful tool for determining the optimal sizes and spacing of the aerators.

5.5.1. Oxygen Transfer Efficiency

The oxygen transfer efficiency of an aeration installation, in a flowing stream, can be approximated by measuring the rate of stream flow and the difference in DO levels upstream and downstream of the aerator. The profile of the DO levels in the vicinity of the aerator should be determined, because it gives the low and the high DO levels upstream and downstream of the aerator, respectively. Neglecting the effects of other oxygen sources and sinks, the steady-state equation for estimating the oxygen transfer efficiency of the aerator is as follows:

$$N = \frac{\mathrm{b}Q(C_\mathrm{d} - C_\mathrm{u})}{P} \tag{77}$$

where N is the oxygen transfer efficiency of aerator, b is the conversion factor, Q is the rate of stream flow, C_d and C_u are, respectively, the DO concentrations at downstream and upstream points, and P is the brake horsepower of aerator.

When estimating the value of oxygen transfer efficiency, N, for the standard zero DO level from a field test result, the following equation can be applied:

$$N_\mathrm{o} = \frac{NC_{\mathrm{S(20°C)}}}{C_{\mathrm{S(T)}} - C_\mathrm{m}} \tag{78}$$

where N_o is the oxygen transfer efficiency of aerator at zero DO level, $C_{\mathrm{S(T)}}$ is the saturation oxygen concentration at temperature T, C_m is the DO concentration at the aerator, and it can be approximated by an average value of upstream DO level, C_u, and downstream DO level, C_d, or by a logarithmic average based on the basic aeration equation (57, 58). $C_{\mathrm{S(T)}}$ is the saturation oxygen concentration at water temperature T(°C).

The rate of oxygen transfer shall be further corrected for the differences in the temperature, atmospheric pressure, and river water quality, as described earlier in this chapter. It should be noted that the oxygen transfer efficiency of an instream aerator, N, will vary with the depth and the width of the flow, and hence with the rate of the flow. The relationship between the oxygen transfer efficiency and the rate of stream flow must be known in predicting the oxygen transfer efficiency of an aerator for a different flow condition. When selecting an aeration system for an instream aeration application, the complex interrelationships between the various factors that affect the oxygen balance of the stream should be fully analyzed.

5.5.2. Instream Aeration Systems

Various types of mechanical surface aerators, diffusers, side-stream mixing systems, down flow contactors are available for the purpose of instream aeration. They are described briefly in the following.

5.5.2.1. Mechanical Surface Aerators

A variety of surface aerator devices are available, mostly used in waste treatment applications. For applications in rivers and streams, the device is required to be float-mounted so that it can be operated under a wide range of flow conditions. Because the stream current will contribute in providing the aerator with a supply of water low in DO to the aerator mixing zone, an aerator that expends relatively low pumping energy can be more economical. The width of induced aeration zone of an aerator, a design parameter for determining the number of units required over a stream section, is reported as varying with the aerator sizes: 300 to 400 ft for 100-hp units and out to 150 ft for 10 to 15-hp units (59). Based on analytical solutions, the diameter of the zone is found limited to approximately 4 to 10 times the water depth (60). The average oxygen transfer rate under standard conditions is found ranging from 1.5 lb O_2/hp-h to 4.5 lb O_2/hp-h with higher transfer rate being associated with high flows. The increase in oxygen transfer efficiency results from the longer detention time for bubbles, generated by the greater stream velocity, which also caused a shearing effect on the bubbles (29, 56).

5.5.2.2. Diffuser Aerators

In diffuser systems, air or molecular oxygen is piped to a distribution system where it is introduced to the water through nozzles, orifices, or jets. The diffuser heads can be installed at various depths below the water surfaces. Note at each depth a set of diffusers heads must have an isolation valve so that set of diffuser head can only be operated. Increased height of water provides greater contact time and thus increased oxygen absorption, but also increases the hydrostatic head of air pumping. When using oxygen instead of air, the head is not a problem as the source would be under high pressure. Other parameters such as bubble sizes, air-flow rates, and water velocities are factors affecting the efficiency of the diffuser system. A diffuser aerator may be less economical than a mechanical aerator, but the flexibility of diffused aeration must be considered when making comparisons.

5.5.2.3. Downflow Contactors (U-Tube Aerators)

Aeration is accomplished by temporarily pressurizing an air-water mixture as it is forced downward by a slight head over a vertical tube (Figure 5.6). More oxygen is transferred near the bottom of the tube because of increased pressure at lower temperature, which also increases the DO deficiency. Several different types of U-tube systems can be designed: (a) air or oxygen is injected into the inlet water by a blower, (b) air-water mixture is provided by a cascade, and (c) air is introduced by a venturi differential that is vented to the atmosphere (29, 61–64).

The transfer of oxygen to water is more effective the deeper the U-tube. However, the saturation of dissolved nitrogen also increases, a critical factor in design because fish are adversely affected when nitrogen supersaturates. A higher air-water ratio also increases the

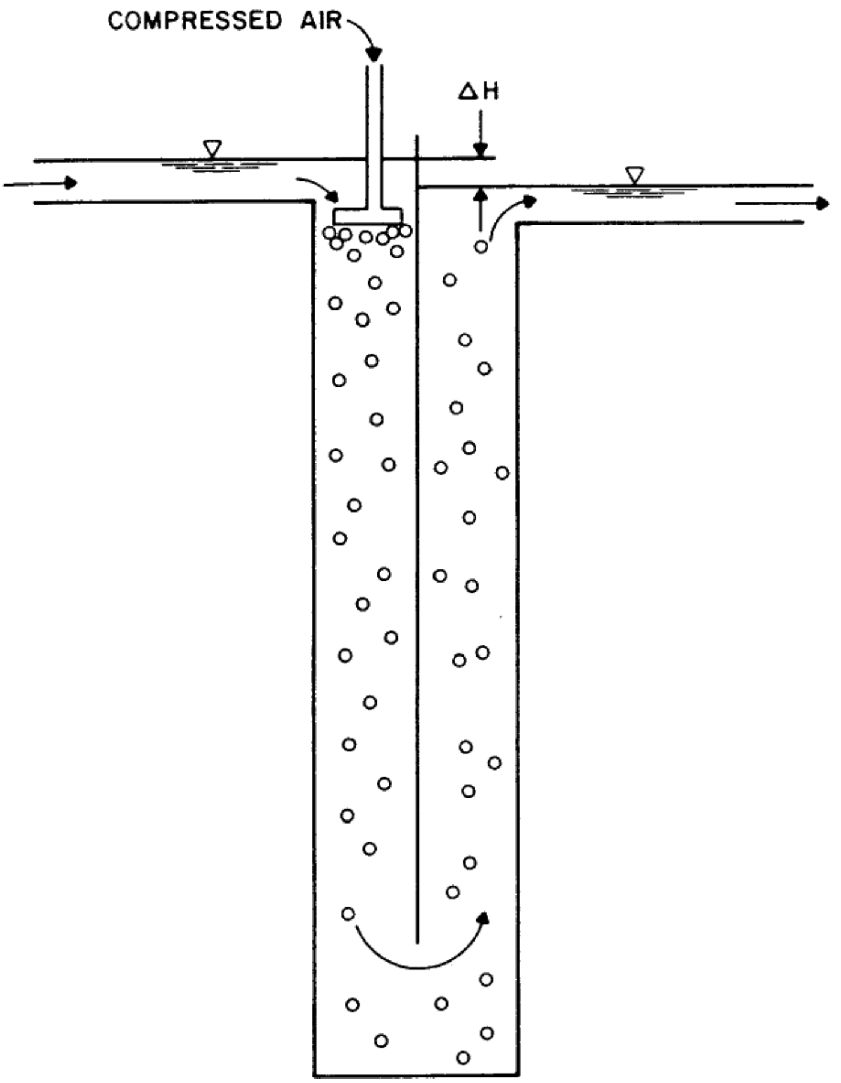

Fig. 5.6. U-tube type aeration system.

oxygen transfer rate; however, if the ratio exceeds that for nitrogen saturation (generally 10% to 20% higher than that required for oxygen saturation), then nitrogen supersaturation will also occur.

Use of pure oxygen in place of air is a method of avoiding nitrogen supersaturation. An oxygen injection system yields the change in DO approximately five times that of an air injection. Lower velocity through the U-tube is found to give more efficient oxygen transfer. However, the use of smaller U-tubes with higher velocities may be more economical if one is interested only in transferring a given amount of oxygen.

5.5.2.4. Sidestream Pressurization

A small percentage of the flow volume is drawn off, mixed with oxygen under pressure, and the resulting supersaturated mixture diffused back into the river. This method requires the construction of an on-site gaseous-oxygen generating plant, and thus the cost of oxygen is a major factor in its design. The rate of oxygen transfer increases with the increasing oxygen content of the gas, and if pure oxygen is used for aeration, the rate is independent of the DO content in the water between 0 to 12 mg/L (65, 66).

5.5.2.5. Reaeration at Dams, Weirs, and Cascades

The reaeration occurring at dams, weirs, and cascades is caused by the highly turbulent flow at the base of these structures, generates new water surfaces and hence increases the oxygen transfer rate. The change in DO concentration depends on the upstream DO values and the height of free-fall, and the majority of the oxygen transfer takes place at the splash area and not in the falling water.

Although the reduction in deficit over these structures may be significant, reduced by about one-half over a free-fall of 5 to 6 ft, the overall effect of the structure may be detrimental to the stream reaeration capacity. Decreased velocity and natural turbulence and increased depth are the result of a flow-retarding structure.

It is possible that the decrease in the natural reaeration (K_2) is greater than the increase from artificial aeration caused by the free fall (67–69).

6. SPRAY AERATION

6.1. Introduction

Spray aeration has its greatest application in the field of water treatment. The method accomplished the gas transfer process by causing the water to break into drops or thin layers, thereby increasing the area-volume ratio and the gas transfer process. The spray aeration method includes spray nozzles, cascades, and multiple trays. The applications of aeration for the improvement of water supply quality are summarized as follows:

1. Removal of tastes and odors originating from:
   a. Essential oils of algae and other living organisms.
   b. Decomposition of organic matter.
   c. Hydrogen sulfide.
   d. Chlorination of water.
   e. Iron and manganese that produce a metallic or chalybeate taste.
2. Removal of dissolved gases:
   a. CO_2, naturally present from aerobic or anaerobic decomposition of organic matter liberated by alum.
   b. H_2S, like CO_2, anaerobic decomposition product.
   c. Cl_2, in excess of that needed for disinfection.
   d. SO_2, excess after dechlorination
3. Changing the pH:
   a. By removing CO_2 to reduce the corrosiveness.

4. Addition of gases:
   a. Oxygen, to distilled water to remove flat taste in deferrization or demanganization.
   b. CO_2, for recarbonation of softened waters and for assistance in H_2S removal.

6.2. Types of Spray Aerators

6.2.1. Spray Aerators (Nozzle Aerators)

Most spray aerators take the form of pipes equipped with nozzles that spray water into small droplets, thus a large area-volume ratio can be produced. Gas transfer efficiency of nozzled aerators is relatively high. In general, the design and the selection of nozzle types should consider factors such as ensuring a minimum of clogging, easy cleaning, no moving parts, and so on.

Many types of nozzles are available. Water may be sprayed upward, horizontally, or downward. The following types of nozzles, which are sketched in Figure 5.7, are briefly discussed.

1. Berlin nozzle: two jets of water impinge at an angle of 90° and are dispersed.
2. West Palm Beach nozzle: by means of a tangential inlet in the head of the nozzle, a revolving flow of water is obtained, resulting in a fine uniform spray distribution.
3. Sacramento nozzle: a movable central cone above a bell mouth produces a thin film bending somewhat upward and then disintegrating into droplets.
4. New York nozzle: vanes inside the nozzle are such that the water film flowing out from the discharge orifice is fan shaped.
5. Oblique nozzle: by means of obliquely bored channels, a rotating movement of the water is obtained.
6. Dresden nozzle: the water flowing out of a copper tube strikes a circular glass disc and is stretched into a thin circular film, disintegrating into small droplets.

Of the foregoing nozzles discussed, the Dresden type seems to have the most advantages in that (a) the inlet head may be reduced to practically zero without affecting the gas exchange, (b) cleaning is very simple and does not take up much time, (c) the cost and the performance are more favorable than the others, (d) the construction allows more flexible variations of the diameter and the length of the tube, the diameter of the disc, the distance between the tube and the disc. The dimensions described in Table 5.3 are given as the best designs.

Nozzle spray aerators are efficient in oxygen transfer and carbon dioxide removal. Some disadvantages are that they (a) require a large area, (b) can pick up airborne contaminants. (c) can be noisy in residential areas, and (d) have freezing problems in cold climate areas.

6.2.2. Multiple-Tray Aerators

Multiple-tray aerators generally consist of a distribution pan or a perforated pipe grid, multiple levels of slated or perforated trays filled with cokes or gravel, a collecting pan, an enclosure, and an induced or forced draft ventilation system. Air flows countercurrently with the falling droplets of water in the aerator. Multiple-tray aerators have a wide application in aeration of ground water for iron and manganese removal and carbon dioxide gas removal.

Table 5.3
Design data of dresden nozzles

Tube diameter, mm	Length of tube, cm	Disk diameter, cm	Distance between tube and disk, cm
12	15	3	2
19	20	4	3
25	20–25	6	3
41	30–35	6–9	3

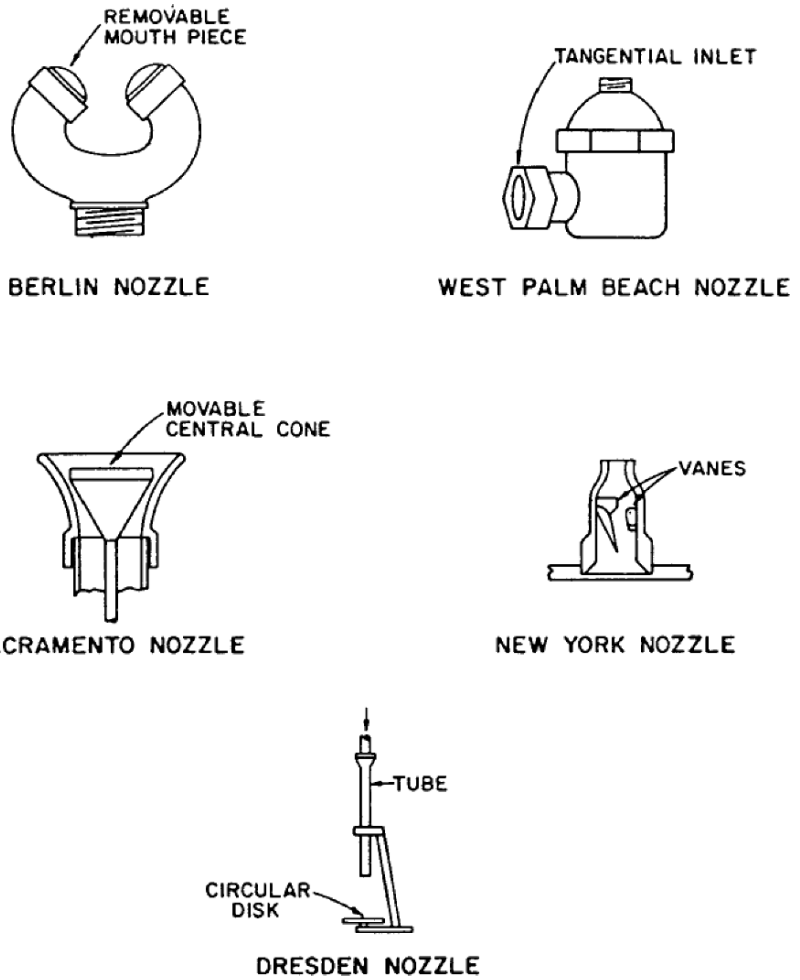

Fig. 5.7. Types of aerator nozzles.

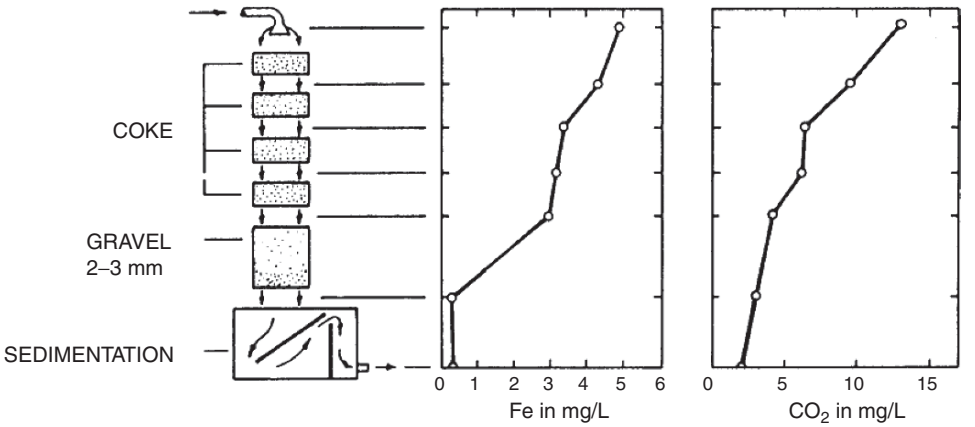

Fig. 5.8. Multiple-tray aerator and typical CO_2 and Fe removals effects.

An empirical equation for estimating the removal of carbon dioxide by multiple-tray aerator, developed by Scott (70), is

$$P = 100\, e^{-kn} \tag{79}$$

where n is the number of trays, k is a constant ranging from 0.12 to 0.16, and P is the percent of free CO_2 remaining after aeration. A sketch of the multiple-tray aerator is given in Figure 5.8, in which the effects of the aeration on carbon dioxide and iron removal are also shown. The carbon dioxide content is reduced from 13 to about 2 parts per million (ppm); the greatest reduction is accomplished by the spray aerator (a distribution pan) and the first tray. The iron content is reduced from 3 to 5 ppm in passing through the coke trays, while the fine gravel trays below provide a very good filtering effect. The effluent only contains about 0.2 ppm of Fe.

Many disadvantages of multiple tray aerators are: (a) carbon dioxide removal will vary seasonally and with changing wind conditions, (b) they show a tendency to clog when the water contains high levels of iron and hinder the trickling of finely divided water droplets, (c) high expenses for cleaning or replacing the cokes. To cope with these disadvantages, forced or induced draft aerators are now more common than the atmospheric-type aerators, and backwashing equipment is used in new aerators to remove the hydrates.

6.2.3. Multiple-Tray Aerators

Water falls and splashes over a series of steps or trays can result in effective aeration. Aeration is mainly achieved by the mixing of air with the falling water in the underlaying steps. The creation of turbulence in water is important, because better aeration results are obtained by increasing the water flow to an optimal rate. Cascade aerators do not require inlet heads, relatively large quantities of water can be treated in a comparatively small area, and they are easy to clean. The structure is simple and inexpensive.

The removal of carbon dioxide by cascade aeration is less efficient in comparison with other aeration methods. In the transfer of oxygen, there is no substantial difference from others. Thus, if raising the oxygen content of the water is the main purpose of aeration, cascade aerators will be very suitable. A reduction of the oxygen deficit by about 30% per step of cascade aerator can be obtained. The aeration of water falling over weirs or dams in streams is a form of cascade aeration.

6.3. Spray Aeration Applications

6.3.1. Hydrogen Sulfide Removal

Hydrogen sulfide gives off unpleasant odor even at very low concentrations. Its high chlorine demand is undesirable and it is also responsible for the destruction of cement and concrete and for the corrosion of metals. Hydrogen sulfide can be removed from water by aeration. Its removal is effective when the water is low in pH.

The solubility of H_2S is about three times that of CO_2. Therefore, when water is aerated, the CO_2 is removed much more rapidly than the H_2S. Reduction of CO_2 content in water raises the pH of the water, thereby causing the H_2S ionization equilibrium to move toward the formation of more sulfide (HS^- and S^{2-}) that cannot be removed by aeration. Therefore, to increase the efficiency of H_2S removal by aeration, the removal of CO_2 or the raising of the pH of the water must be controlled. The aeration of the water in an atmosphere containing a high partial pressure of carbon dioxide is one method of achieving it: A very high concentration or partial pressure of CO_2, of the order of 10% compared to about 0.03% in ordinary air, can be maintained by recarbonation equipment. The result is that the CO_2 content in the water increases and the pH is lowered and maintained at a low value during the aeration. This condition favors the H_2S form and transfer of H_2S is affected. The CO_2 added can be removed in a second aerator of conventional design.

The chemical equation shows CO_2 content below the equilibrium will result in decomposition of the calcium bicarbonate and deposition of the insoluble carbonate. Water containing an excess of CO_2 will change the carbonate into bicarbonates. That is, the excess carbon dioxide will attack the calcium carbonate present in concrete and mortar. This excess carbon dioxide is also corrosive to metals if the alkalinity is low as shown in Figure 5.9. In most aeration systems the residual carbon dioxide concentration is about 2.5 mg/L or more. As a 2.5 mg/L CO_2 corresponds with an equilibrium concentration of 120 mg/L bicarbonate, thus in soft water (low alkalinity) the carbon dioxide content has to be removed further by chemical treatment.

Obviously, the equilibrium concentration of CO_2 can be lowered by maintaining as low partial pressure as possible in enclosed aerators, such that the release of CO_2 can be facilitated. However, even though the CO_2 released is vented, the partial pressure of CO_2 in the enclosed aerator would increase as fast as CO_2 removal occurs, thus slowing down the rate of CO_2 release. Therefore, aeration under pressure in a closed aerator is inefficient for CO_2 removal.

6.3.2. Carbon Dioxide Removal

Most groundwater contains an excess amount of carbon dioxide that has to be removed in water treatment. The equilibrium concentration of CO_2 in water is about 0.5 mg/L. Therefore,

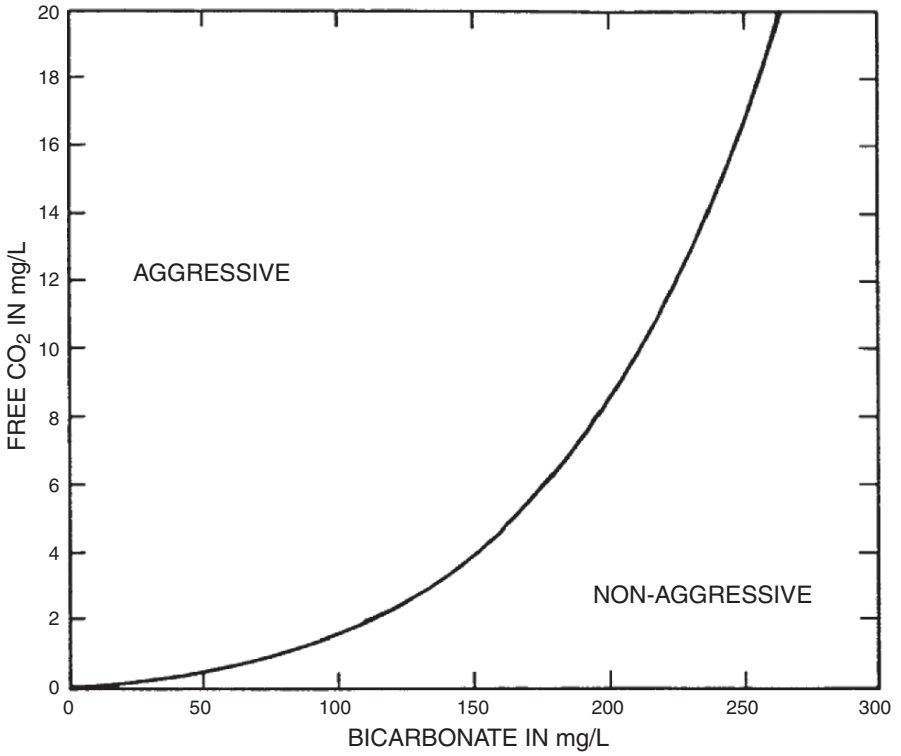

Fig. 5.9. The free carbon dioxide–bicarbonate equilibrium curve.

aeration will not reduce the concentration of CO_2 below 0.5 mg/L. Complete removal of CO_2 requires the use of an alkali such as soda ash or lime. Carbon dioxide in water is in equilibrium with carbonic acid and the bicarbonate as

$$Ca(HCO_3)_2 \Leftrightarrow CaCO_3 + CO_2 + H_2O \tag{80}$$

Because most iron-bearing waters also contain an excess of carbon dioxide, aeration of the water not only oxidizes iron, but also expels the carbon dioxide.

6.3.3. Deferrization and Demanganization

Assuming that the iron in groundwater water is present as ferrous bicarbonate, then the deferrization process proceeds as:

$$4Fe(HCO_3)_2 + O_2 + 2H_2O \rightarrow 4Fe(OH)_2 + 8CO_2 \tag{81}$$

The aeration of the water result in the formation of ferric hydroxide in the water. This formation raise the pH value of the water, which in turn promotes the settling of the precipitated iron oxide. However, the carbon dioxide generated by the above chemical reaction is not removed. A second aeration is needed, or lime is added, to remove the rest of the carbon

dioxide. In general, iron removal in hard water is less difficult than in soft water, as can be seen from the above chemical equation.

Equation (81) also indicates that little aeration is required for the deferrization, where only 0.14 mg/L of O_2 is needed to convert 1 mg/L of Fe^{2+} to $Fe(OH)_2$ (ferric hydroxide). Vigorous aeration of hard water may remove more than the excess CO_2 and cause to precipitate finely divided $CaCO_3$ [Equation (80)]. At pH values above 7.1, the $CaCO_3$ precipitates are negatively charged, may be adsorbed on the positively charged ferric hydroxide particles and remain in colloidal suspension, and require additional treatment such as coagulation for their removal.

Most of the iron-bearing waters also contain manganese. However, demanganization requires more intense oxidation than iron removal. If the manganese content is low, it will be removed together with the iron by aeration. If the manganese content is high, only a slight decrease in the manganese content is effected, but it will be a complete or nearly complete deferrization. Renewed aeration is needed to give manganese-free water.

6.3.4. Removal of Tastes and Odors

Aeration is an effective method for the removal of tastes and odors, including substances such as the volatile and essential oils produced by algae. The tastes and odors caused by industrial wastes such as phenol cannot be effectively removed by aeration alone. The difficulty is that most of these compounds are either high in solubility or very low in vapor pressure. The latter is an important factor in the case of phenol, which has a fairly high boiling point. High boiling point substances have lower vapor pressures at a given temperature than a low boiling point substance. The vapor pressure of phenol is almost negligible, so that it would not be at all practical to attempt to remove it by aeration. Water temperatures usually vary from 0 to 30°C. Within this temperature range, substances that boil at much more than 0°C probably cannot be removed by aeration unless their solubility is quite low.

Chlorine has a boiling point of about −34°C, but it can not be removed readily by aeration because of its fairly high solubility and its tendency to form compounds with water or other adventitious substance. The above discussion does not include the possible subsequent oxidation of these compounds by dissolved oxygen added by aeration. Certain taste and odor producing substances may be changed chemically by the addition of dissolved oxygen.

6.4. Spray Aerator Design

The design of spray aerators involves mainly the hydraulic computation of the jet spray of water through the nozzles, the spacing or arrangement of the nozzles, and the nozzled piping system. The nozzle type is generally selected on the basis of lower cost, minimal maintenance, and high hydraulic efficiency (i.e., high coefficients of discharge and velocity). As shown in Figure 5.10, the hydraulics of a nozzle with a deflecting cone can be expressed by the following equations, if air resistance is neglected:

Velocity of spray:

$$v = c_v\sqrt{2gh} \tag{82}$$

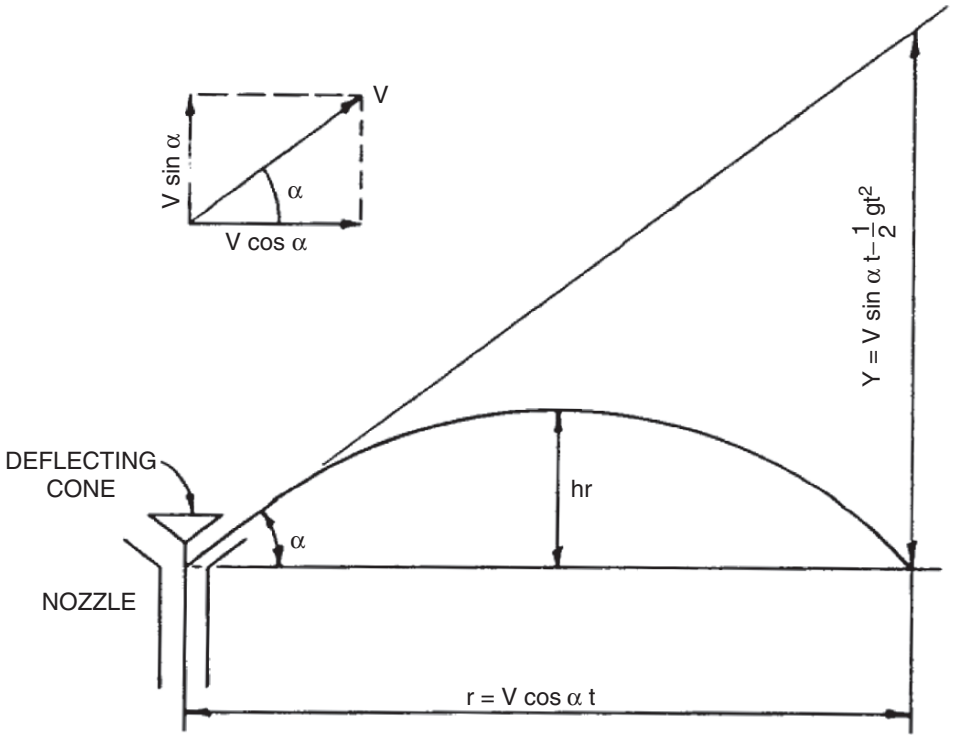

Fig. 5.10. Hydraulics of a spray nozzle.

Time of exposure:

$$t = 2c_{\mathrm{v}}\sqrt{2\frac{2h}{g}}\sin\alpha \tag{83}$$

Radius of spray:

$$r = 2c_{\mathrm{v}}^2\, h \sin 2\alpha \tag{84}$$

Rise of spray:

$$h_{\mathrm{r}} = \frac{1}{2}c_{\mathrm{v}}^2\, h(1 - \cos 2\alpha) \tag{85}$$

Discharge of the nozzle:

$$q = \mathrm{ca}\sqrt{2\mathrm{gh}} \tag{86}$$

where v is the velocity of spray issuing from the nozzle, h is the head on the nozzle, α is the angle of the deflector (or the nozzle) from the horizontal, g is the gravity constant. a is the area of the nozzle orifice, and q, the rate of discharge, t is the time of exposure, h_{r} is the height of

Fig. 5.11. Air resistance in droplet aeration.

rise, and r is the radius or horizontal carry of spray, and c and c_v are coefficients of discharge and velocity, respectively.

The effect of air resistance, which has not been considered in the above equations, is: (a) the radius of spray will be about half the value computed by Equation (84); thus, in designing nozzle spacing, a test should be run on the nozzle to obtain value of r, and (b) the time of exposure will be very little effected because, although the time required to obtain maximum height will be decreased, the time required to fan should be increased by almost the same amount. Figure describes the effect of air resistance to a droplet, as a function of $v/\sqrt{gd}$, where d is the diameter of the droplet. Note that the influence on the height is about twice that on the time (of rise).

6.4.1. Example 1

Assuming a velocity coefficient, c_v, of 0.85 for a well-rounded nozzled central fountain, as shown in Figure 5.12, operating under an effective head of 10 ft, find (a) the rate of discharge, (b) the maximum vertical rise of the spray, (c) the time of exposure, (d) the maximum radius of the spray, and (e) the head required if the discharge is 5 MGD (7.74 cfs).

Solution

Area of orifices:

1–3″ diameter	$7.07\,\text{in}^2$
12–$1\frac{1}{4}$″ diameter	$14.73\,\text{in}^2$
24–7/8″ diameter	$14.43\,\text{in}^2$
Total area	$36.23\,\text{in}^2$ or $0.252\,\text{ft}^2$

(a) By Equation (86):

$$q = (0.85)(0.252)\sqrt{2(32.2)(10)} = 5.44\,\text{cfs}$$

(b) By Equation (85):

$$h_r = \frac{1}{2}(0.85)^2(10)[1 - \cos(120^\circ)] = 5.42\,\text{ft}^2$$

(c) By Equation (83):

$$t = 2(0.85)\sqrt{\frac{2(10)}{32.2}}\sin(60^\circ) = 1.16\,\text{s}$$

(d) By Equation (84):

$$r = 2(0.85)^2(10)\sin(120^\circ) = 12.5\,\text{ft}$$

(e)

$$v = \frac{Q}{A} = \frac{7.74}{0.252} = 30.7\,\text{fps}$$

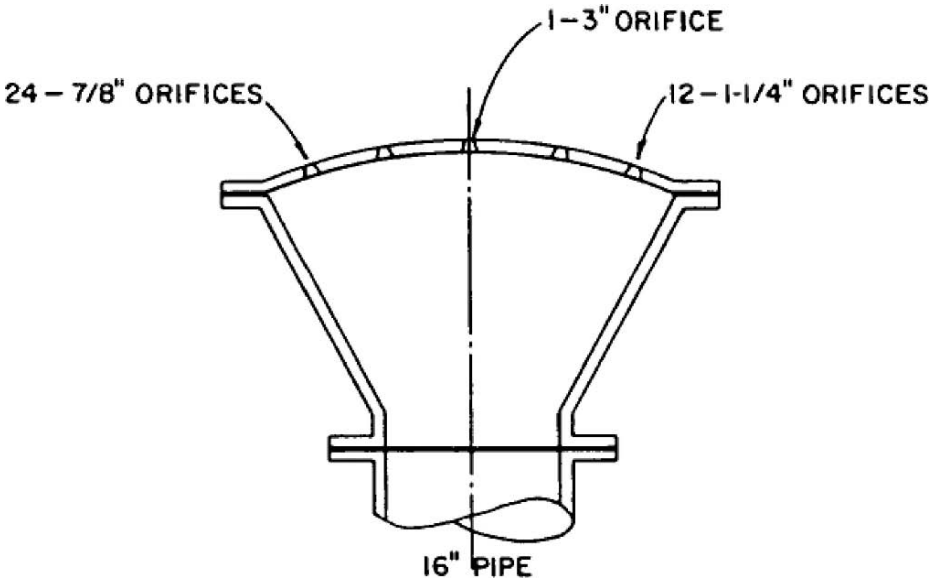

Fig. 5.12. Detail of central fountain spray aerator.

Head required to create the velocity in orifices,

$$h = \frac{v^2}{2gc_v^2} = \frac{(30.7)^2}{(64.4)(0.85)^2} = 20.3\,\text{ft}$$

Or by the ratio:

$$h = h_o\left(\frac{Q}{Q_o}\right)^2 = 10\left(\frac{7.74}{5.44}\right)^2 = 20.2\,\text{ft}$$

The piping for the nozzles should be designed so that all nozzles discharge nearly the same amount of flow. Equity of discharge can be achieved by controlling the loss of head in the piping to the small in comparison to the losses in the orifices. In practice, the head loss or the operating head in the orifices is set between 3 and 15 ft and the permissible friction loss in the piping for the nozzles is about 10% to 20% of that in the nozzle.

If the discharge from the farthest orifice in the piping is to be held to a value equal to mq_1, where q_1 is the discharge from the first orifice and m is the ratio of flow from the farthest orifice q_2 to q_1, then, based on the common orifice formula $q = \text{ca}\sqrt{2\text{gh}}$, the head in the farthest orifice, h_2, is

$$h_2 = \frac{h_1 q_2^2}{q_1^2} = m^2 h_1 \tag{87}$$

Because $h_2 = h_1 - h_f$, where h_f is the friction loss in the piping, the permissible head loss h_f is

$$h_f = (1 - m^2)h_1 \tag{88}$$

For a $m = 0.9$, for example, $h_f = 0.19\,h_1$

If the diameter of the piping for the nozzles is kept constant, the friction loss from the entrance of the piping to the farthest orifice is approximately equal to the loss of head caused by the entrant flow (i.e., total flow) passing through 1/3 of the length of the piping, or

$$h_f = \frac{1}{3} S_o L \tag{89}$$

where S_o is the slope of the hydraulic gradient at the entrance, and L is the length of the piping. Substituting Equation (89) into Equation (88), obtain the permissible hydraulic slope as:

$$S_o = 3(1 - m^2)\frac{h_1}{L} \tag{90}$$

6.4.2. Example 2

Design an aeration lateral for nozzles of 30 feet in length, operating under an effective head of 9 feet in the entrant manifold with a flow rate of 5 cubic feet per second (cfs).

Solution

Assuming the nozzle to be used has an coefficient of velocity, $c_v = 0.92$
Velocity in the entrant nozzle,

$$v = c_v\sqrt{2\ gh} = 0.92\sqrt{2(32.2)(9)} = 22.2\,\text{fps}$$

Required nozzle area,

$$A = \frac{Q}{v} = \frac{5}{22.2} = 0.225\,\text{ft}^2 = 32.4\,\text{in.}^2$$

Use 22 to $1\frac{3}{8}''$ nozzles ($1.4849\,\text{in.}^2$/nozzle)
Selecting a $12''$ diameter pipe.

$$\text{Velocity in } 12'' \text{ pipe} = \frac{5}{\pi(1)^2/4} = 6.36\,\text{fps}$$

Friction loss in a $12''$ diameter pipe is determined from the Hazen-Williams as presented in Equation (91) with a C value of 100.

$$h_f = \frac{4.73\text{Q}^{1.85}\text{L}}{\text{C}^{1.85}\text{D}^{4.87}} \tag{91}$$

$$h_f = \frac{4.73(5.0)^{1.85}(30)}{(100)^{1.85}(1.0)^{4.87}} = 0.56\,\text{ft}$$

By Equation (90), and solving for m

$$S_o = 3(1 - m^2)\frac{h_1}{L} \tag{90}$$

$$0.56 = 3(1 - m^2)9/30$$

$$m = 0.91$$

That is,

$$q_2 = 0.91\ q_1$$

7. RECENT DEVELOPMENT IN SURFACE AND SPRAY AERATION

To reduce energy cost and improved treatment, researchers (72) have examined the methods to control the speed of surface aerators. Other researchers (73, 74) presented a method of improving the amount of oxygen delivered by surface aerator to wastewater. U-Tube aeration system (75) and high circulation airlift reactors (76, 77) were shown as methods for providing high oxygen absorption efficiencies.

Researchers (72) developed an automatic control system to control the speed of surface aerators. The system is used to maintain the dissolved oxygen concentration in aeration basins at a constant level. At high organic loading more oxygen is required; the speed of the surface aerator would be increased which increase the amount of oxygen being transferred into the wastewater. At low organic loading less oxygen is required; the speed of the surface aerator would be decreased which decrease the amount of oxygen being transferred into the wastewater. The activated sludge process performance is improved when the amount oxygen supplied to the microorganism is proportional to the organic loading or the organic food supply for the microorganism. A deficient level of oxygen could result in poor performance of an activated sludge process by degrading effluent quality, and causing sludge bulking and proliferation of filamentous bacteria. An excess level of oxygen could cause the destruction of flocs from excessive mixing, which would result in poor settling characteristic of the activated sludge and poor performance in clarification process. To supply excess oxygen to aeration basin results in wasting energy and increases energy cost to operate the wastewater treatment facility. The authors developed a PI autotuning algorithm using an auto-regressive exogenous (ARX) model so that the tuning parameters could be readily obtained. This PI controller adjusts the speeds of surface aerator to maintain a constant dissolved oxygen (DO) level in the aeration.

The researchers examined the DO levels in an aeration basin of a wastewater treatment plant treating coke wastewater and compared them when the PI controller adjusted the speed of the surface aerators to a constant speed operation of the surface aerators. To tune the parameters of the PI controller for the discrete ARX model the authors used a first order plus time delay desired trajectory for the unit step change of the set point. The monitoring included more than 100 input/output signals for control of the DO level in the aeration basin.

Without the controller on the aerator the DO levels ranged from ranged 0 mg/L to 2.0 mg/L, whereas the DO levels with PI controller on the aerator consistently track the set point of

1.0 mg/L. The researchers estimated that savings of 42% in electricity consumption could be realized when the PI controller is used to adjust the speed of a surface aerator over a constant speed surface aerator.

Effects of velocity baffles on the performance of disc aerators were examined in oxidation ditch treating domestic wastewater (73, 74). The velocity baffles are located downstream of the surface disc aerators. The baffles were used to direct highly aerated surface wastewater towards the bottom of the channels.

The oxidation ditch consists of inner and outer channels with volumes of 0.335 and 0.716 million gallons, respectively. The influent enters in the outer channel and effluent discharges from the inner channel. Operational water depth was approximately 12 ft.

The inner channel was aerated with two disc surface aerators located 180 degrees apart. The outer channel was aerated with three disc surface aerators located at 90 degrees, 180 degrees and 270 degrees around the outer channel. Each aerator used a 40 horsepower motor and contained 40 disc operated with 18 in. immersion. Test was conducted with one inner aerator operated at 44 rpm and one at 60 rpm, while the outer channel was operated with one aerator at 60 rpm and two at 53 rpm. Manufacturer's reported Standard Oxygen Transfer Rates (SOR) per disc operated at 60 rpm is 2.85 lb/h, for disc operated at 53 rpm is 2.36 lb/h and for disc operated at 44 rpm is 1.73 lb/h. Using immersion correction factor of 0.87 for 18" immersion, the SOR for aerators under test conditions was approximately 423 lb/h.

The authors measured and examined velocities, dissolved oxygen levels and the oxygen uptake rates in each channels. Measurements were conducted with baffles full up, 1/3 immersed, 2/3 immersed and fully immersed positions. The velocities and dissolved oxygen levels were measured 3, 6 and 9 feet below water surface. Velocities were measured downstream of each aerator at the inner edge, center and outer edge of the channel. Oxygen levels were measured upstream on each aerator.

Tables 5.4, 5.5 and 5.6 present the average channel velocities, the average channel dissolved oxygen levels and oxygen delivered to the wastewater, respectively. During the eleven day test period, the average flow was 4.28 MGD and the $CBOD_5$ removed was 197 mg/L. The volume weight average velocity with baffle fully up was found to be 2.52 ft/s and with baffle in the

Table 5.4
Average channel velocities

Baffle position	Channel	Average velocity ft/s	Volume weight average velocity ft/s
Fully Up	Outer	2.82	
	Inner	1.88	2.52
1/3 Immersed	Outer	2.80	
	Inner	2.02	2.55
2/3 Immersed	Outer	2.23	
	Inner	1.79	2.09
Fully Immersed	Outer	1.68	
	Inner	1.07	1.49

Table 5.5
Average channel oxygen levels

Baffle position	Inner channel mg/L	Outer channel mg/L
Fully up	0.52	0.33
1/3 Immersed	1.21	0.36
2/3 Immersed	0.50	0.26
Fully immersed	0.52	0.25

Table 5.6
Oxygen delivered to the wastewater

Baffle Position	Channel	OUR mg/L/hr	Oxygen delivered lb O_2/hr
Fully up	Outer	54.5	326
	Inner	28.5	80
	Total		406
1/3 Immersed	Outer	54.3	325
	Inner	40.1	112
	Total		437
2/3 Immersed	Outer	70.7	442
	Inner	51.2	144
	Total		586
Fully Immersed	Outer	81.7	488
	Inner	60.4	169
	Total		657

fully immersed position velocity was found to be 1.49 ft/s. The average DO levels with baffles in fully up position were 0.52 mg/L for the outer channel and 0.33 mg/L in the inner channel and the levels with baffles in fully immersed position were 0.52 mg/L for the outer channel and 0.25 mg/L in the inner channel. The total amounts of oxygen delivered to the basin were 406 lb-O_2/h with baffles fully up position and 657 lb-O_2/h with fully immersed baffles.

The average velocities and average dissolved oxygen levels in the channels decreased when the baffles were fully immersed compared to baffles in the fully up position (no baffles). The position of the baffles has no effect on the oxygen demands in the channel, since demand is dependent on the daily $CBOD_5$ loading to the basin. The daily $CBOD_5$ loadings to the basin were similar for the test days with only the usual fluctuations throughout day. The oxygen uptake rate (OUR) measures the oxygen consumed by the microbial population expressed in mass of oxygen consumed over time per unit volume. The actual oxygen transfer rate can be approximated by multiplying the OUR by the system volume. The authors concluded that because both the daily organic loadings and the speeds of the aerators did not change, more oxygen was delivered and consumed by the microbial population with baffles in the fully immersed position than in the fully up position (no baffle). As shown in Table 5.6, the amount

of oxygen delivered to the wastewater was increased from 406 lb/hr with baffles in the fully up position (no baffles) to 657 lb/h with baffles in the fully immersed position.

Speece (75) examined the use of U-Tubes with oxygen to supplement the dissolved oxygen level of the Tombigbee River in Demopolis, Alabama, USA. This river receives effluent from two paper mill activated sludge wastewater treatment plants with flows of 35 MGD and 120 MGD that achieve over 90% BOD removal.

Normally the flow in river is largest enough that effluent from these wastewater treatment plants has negligible impact on the river's dissolved oxygen level. During prolong drought and high temperature periods which result in lower river flow and higher river water temperature, the discharge from the wastewater treatment plants are restricted to maintain dissolved oxygen level in the river. The discharge restriction results in the paper mills curtailing their operations. As result of this restriction, methods were examined to supplement the dissolved oxygen level in the river with aeration system.

The author examined the oxygen sources and found that air as oxygen source has two distinct disadvantages when compare to pure oxygen: 1) air content only 21% oxygen and 2) the nitrogen content in the air can cause adverse condition in water when water is aerated with diffused aeration system at high rates. The water can become supersaturated with nitrogen. When water that is supersaturated with nitrogen, it can impair the health fish and even cause death of the fish. Owing to these disadvantages of air, commercial oxygen was selected to be used with a U-Tube aeration system for supplementing the oxygen level in the river.

The U-Tube aeration system can be operated at high pressures, which can increase saturation concentration of water to 250 mg/L at six atmospheres with 100% pure oxygen. This oxygen saturation concentration significantly high than 9.2 mg/L that can be achieved when water is aerated with air with 21% oxygen at one atmosphere.

The State of Alabama required 4.7 kg of DO be add to the river for 1.0 kg of BOD_5 added to the river from paper mill wastewater treatment plants. Additionally, the State limits the DO level of the river to the 100% air saturation concentration for river. At this limit and with a river temperature of 28°C, the DO could not exceed 7.9 mg/L.

A U-Tube aeration system was designed, installed and tested. It was 4 ft in diameter and 175 ft deep that provide a volume of $2,200\ ft^3$. The water flows down the inner pipe and back up to the top through the annular space between the inner and outer pipes. The water velocity of approximately 10 ft/sec greatly exceeds the 1 ft/s buoyant velocity of oxygen bubbles. As a result, as the oxygen gas is introduced at the inlet of the U-Tube, the bubbles are dragged along with the water flow down to bottom of the inner pipe and up through the annular space to the discharge at top of the U-Tube. The oxygen enrich effluent pass through a bubble harvester where undissolved bubbles are collected and return to the influent of the U-Tube. The discharge from U-Tube discharges at the bottom of the river, 35 ft below the water surface through a multiport diffuser.

Testing determined that the discharge from the U-Tube had DO concentrations up 50 mg/L at water temperature of 28°C and provided 14,600 lb of oxygen to the river. Because the DO concentration is less than saturation concentration for pure oxygen at 35 ft of pressure there is no potential for effervescent loss of DO. Additionally, the high DO discharge was mixed throughout the river cross-section within a short distance of the multi-port diffuser. It was

determine that oxygen absorption efficiency for this U-Tube aeration system ranged from 80 to 90%. It was also determined, at 35 MGD flow through the U-Tube, the head loss for the U-Tube was 5 ft. With this head loss plus the 175 ft of static head, the U-Tube aeration system required 1,378 hp to operate.

Hudson et al. (76) and Hink et al. (77) presented papers on high recirculation airlift reactors, which have high oxygen transfer concentration (3 $kg/hr/m^3$) and require much less land than a conventional activated sludge reactor (up 80% less). These reactors consist of head tank on top of pipe ranging in length for 50 to 150 m. The pipe is divided into two unequal sections. The small section is called the downcomer and the larger section is called the riser. Both sections have air injection point that compress air can be injected.

Initially, the wastewater in the reactor is stationary, then air is injected into the riser section causing the wastewater to flow from the downcomer section to the riser section and back in the head tank. Once this flow pattern is established air is gradually injected into the downcomer section, while the amount air injected into the riser section is decreased. Operating velocities in high recirculation airlift are typically 1.5 m/s or higher.

As wastewater flow is introduced into the head tank, it is drawn down the downcomer and is rapidly diluted with internal recycled mixed liquor. The wastewater travels down the downcomer, the static pressure increases, which increase the amount of the injected air that is dissolved into wastewater. The wastewater then travel up through the riser section into the header tank. As the static pressure decreases as the wastewater travel up the riser section, the entrained gases in the wastewater are liberated.

Hicks et al. (77) examined a high recirculation airlift reactor that was 150 m deep with a 0.3 m diameter downcomer and a 0.5 m diameter riser. The air injection point in the downcomer was located at depth of 75 m and the injection point in the riser was located at depth that was less. Air was injected into the riser at flow rate of 30 m^3/h. After the maximum wastewater flow rate of 0.23 m^3/s was achieved and maintained, the transferring of air to the downcomer injection point was commenced. The authors observed that 25 m^3/h of injection air was transferred to the downcomer injection point, while remaining 5 m^3/h continued to be injected into the riser. It was also observed that the air transfer rates of 0.08 m^3/s^2 and 0.02 m^3/s^2 allowed the reactor to achieve a stable operating velocity of 0.23 m^3/s in approximately 900 s. Higher transfer rates resulted in unfavorable conditions of decrease velocity in the reactor and reversal of flow in the reactor.

Overall oxygen transfer coefficient was examined by Cascaval and associates (78) for surface aeration system on bioreactor. Other researchers (79) used surface aeration system on pilot-scale manure storage facility.

Cascaval et al. (78) conducted bench-scale study to examine effects on the overall oxygen transfer coefficient for surface aeration (mixing-sparging equipment). The researchers quantified the effects apparent density, specific power input and superficial air velocity on the overall oxygen transfer coefficient. From these results, the researchers proposed mathematical model that showed good agreement with experiment data.

Researchers (79) examined the reduction of volatile fatty acids (VFA), total solids (TS) and total volatile solids (TVS) when an intermittent surface aeration device was tested on a pilot-scale manure storage facility. The result showed that addition of surface aeration device increased the TS reduction from 9.26% to 26.9% and the reduction of TVS from

16.6% to 46.40%. Additionally, VFA removal efficiency increased exponentially from 60% after one week to 98% three months later. Additional applications of the surface aeration and spray aeration for drinking water and wastewater treatment can be found from the literature (80–82).

NOMENCLATURE

A = area across which diffusion is taking place
BOD = biochemical oxygen demand
C = instantaneous concentration
$\langle C \rangle$ = time-averaged concentration
C_d = dissolved oxygen level downstream
C_i = concentration of gas at the gas—liquid interface
C_{kr} = concentration of dissolved tracer gas remaining in the water at time t
C_{kr_o} = concentration of dissolved tracer gas at time $t = 0$
C_L = concentration of gas in the main body of the liquid
C_S = equilibrium or saturation concentration of gas in the liquid
$C_{S(T)}$ = saturation oxygen concentration temperature T
C_{SW} = saturation concentration of dissolved oxygen of sewage
C_u = dissolved oxygen level upstream
c = stream constant based on river channel
c = coefficient of discharge
c_v = coefficient of discharge
D = gas saturation deficit or $D = C_S - C$
d = diameter of drupet
d = diameter of molecule
D_L = coefficient of oxygen molecular diffusion
D_m = coefficient of molecular diffusion or molecular diffusivity
D_m^t = overall mass transfer coefficient
DO = dissolved oxygen concentration
E = voltage of aerator drive unit
Eff = overall efficiency of aerator drive unit
g = gravity constant
H = mean depth of flow
H' = depth above minimum low water stage
h = head on the nozzle
h_f = fraction loss in pipe
h_r = height of rise
I = amperage of aerator drive unit
K_G = overall gas-film coefficient
K_L = overall liquid-film coefficient
$K_L a$ = overall gas transfer coefficient
K_1 = deoxygenation coefficient in stream
K_2 = stream reaeration coefficient

k_g = gas-film coefficient
k_l = liquid-film coefficient
k_S = gas absorption coefficient
L = Length of pipe
L = concentration of organic matter in BOD
L_f = liquid-film thickness or thickness of surface layer from which gas molecules escape to atmosphere
m = ratio of flow from first orifice to the farthest orifice
m = stream constant based on stream depth
m = mass
$\boldsymbol{N}$ = mass flux vector
N = oxygen transfer efficiency of aerator at test or at design conditions
N_o = oxygen transfer efficiency at standard conditions.
n = stream constant based on stream velocity
n = rate of mass transfer per unit area or mass flux
$\langle n \rangle$ = time-averaged rate of mass transfer per unit area
n^t = overall rate of mass transfer per unit area under combined effect of molecular and turbulent diffusions
P = brake horsepower of aerator drive unit
PF = power factor of aerator drive unit
p = partial pressure of gas in the air in contact with the liquid
p_g = partial pressure of gas in the main body of the gas
p_i = partial pressure of gas at the gas-liquid interface
p_s = the partial pressure corresponding to a saturation concentration as defined by Henry's law
q = rate of discharge
R = a source-sink term
r = radius or horizontal carry of spray
r = rate of surface renewal or $r = 1/t_e$ (Hiqbie)
r = average rate of renewal (Dankwerts)
r_o = oxygen uptake rate
S_o = slope of hydraulic gradient
T = temperature in °C
t = time of exposure
t_e = surface age or time of surface renewal
t' = time internal foe gas transfer the interface
U = mean velocity of flow
U' = longitudinal turbulent velocity
V = volume of liquid or system
$\mathbf{V}$ = velocity vector
$\langle V \rangle$ = time-averaged velocity vector
W = total mass (or weight) of gas transferred

dC/dt = rate of gas transfer with time
dC/dy = concentration gradient across the diffusing area
β = oxygen correction factor for waste defined as the ratio of the C of wastewater to the C_S of clean water
α = overall transfer coefficient correction factor for waste define as the ratio of the $K_L a$ of wastewater to the $K_L a$ of clean water
∇ = symbol grad or gradient defined as $\nabla = i\partial/\partial x + j\partial/\partial y + k\partial/\partial z$
∇^2 = Laplace operator defined as $\nabla^2 = \nabla \cdot \nabla = \partial^2/\partial x^2 + \partial^2/\partial y^2 + \partial^2/\partial z^2$
θ = temperature coefficient for liquid-film coefficient
μ_o' = normal velocity of gas molecules to leave the liquid phase
Λ = turbulent scale parameter

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6
Activated Sludge Processes

Lawrence K. Wang, Zucheng Wu, and Nazih K. Shammas

CONTENTS

Abstract Activated sludge consists of suspended biological flocs that are matrices of microorganisms, nonliving organic matter and inorganic materials. The activated sludge or biological flocs mix with the waste stream, oxidize the organic substances in the wastewater in the presence of oxygen for bio-oxidation and nitrification reactions, or in the absence of oxygen for denitrification reaction. This chapter introduces the suspended growth systems, bio-oxidation, microorganisms, substrate removal, enzymatic actions, energy flow, microbial synthesis, respiration, kinetics, sludge growth, complete-mix bioreactor, plug-flow bioreactor, contact stabilization, extended aeration, conventional activated sludge, step aeration, Kraus process, tapered aeration, modified aeration, high-rate aeration, oxidation ditch, pure oxygen activated sludge, flotation activated sludge, and process design.

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Key Words Conventional activated sludge • complete-mix activated sludge • contact stabilization • step aeration • extended aeration • Kraus process • pure oxygen activated sludge • flotation activated sludge • kinetics • process design • examples • cost.

1. CONCEPTS AND PHYSICAL BEHAVIOR

Current practice in the secondary treatment of wastewater calls for the use of biological oxidation to remove organic substances. When it comes to selecting the method of biological oxidation, the pollution control engineer has at his or her disposal a variety of treatment processes, among which activated sludge is currently the most popular. In this section, the function and limitations of the activated sludge processes are reviewed. The principles of biological oxidation and of the energy flow concept are described, and the relationship of synthesis and respiration are discussed in relation to the importance of activated sludge process control.

1.1. Definition of Process

Activated sludge consists of biological flocs that are matrices of microorganisms, nonliving organic matter, and inorganic materials. The microorganisms include bacteria, fungi, protozoa, and higher forms of animals such as rotifers, insect larvae, and worms. An activated sludge process can be defined as a system in which biological flocs are continuously circulated to come into contact and to oxidize the organic substances in the presence of oxygen. The fact that an "active" mass of biological forms is maintained in the system for continuous and successful biological oxidation explains why the process is designated "activated sludge" treatment.

The objectives of activated sludge treatment are twofold: (a) to obtain the maximum possible removal of organic substances with the shortest possible time, and (b) to produce flocculant biological flocs having a good settling characteristic. Both are essential in controlling the secondary effluent quality. From the economic point of view, it is also desirable to meet both objectives because small aeration tank(s) and final clarifier(s) can be used. The two objectives, however, are not compatible. Biological flocs that are very efficient in removing organic Substances at a rapid rate are flocs that normally settle poorly and vise versa. The tradeoff is manifested in the performance of various activated sludge processes. Design engineers and plant operators should be fully aware of the incompatibility of these two objectives for proper design and operation of a plant so that certain specific treatments can be accommodated and optimization of treatment performance can be planned intelligently. For innovative design, the readers are referred to Section 7.3 this chapter, on "Secondary Flotation Process" (1).

It is important to recognize the capability as well as the limitations of activated sludge processes. In the aeration tank biodegradable organics are converted to inorganics. A complete oxidation of organics can be expressed as:

$$\text{Organics (C, H, O, N, P, S)} + O_2 \rightarrow CO_2 + H_2O + NO_3^- + PO_4^{3-} + SO_4^{2-} + H^+$$

The equation above assumes an infinite period of aeration time and plentiful microorganisms needed to carry out the complete oxidation, including nitrosomonas and nitrobacters, are present. Economic constraints do not allow sufficient time for complete oxidation even for an extended aeration process. Nor is it feasible to maintain a steady population of the

less competitive organisms (e.g., nitrosomonas and nitrobacters) in the aeration tank with the present operational scheme. Most noticeably is the lack of nitrification in the process and significant amount of ammonia nitrogen exists in the effluent as result. A typical activated sludge treatment process may yield the following:

Ammonia nitrogen	12 mg/L as N
Phosphate	10 mg/L as PO_4^{3-}
Nitrate	0.1 mg/L as N
Nitrite	0.01 mg/L as N
BOD	20 mg/L
Suspended solids	30 mg/L

Other than high residues of nitrogen and phosphates in the effluent, treatment plant operators are content with a BOD of 20 mg/L and suspended solids of 30 mg/L or thereabouts. The July 1977, effluent guidelines for publicly owned secondary treatment plants issued by the US Environmental Protection Agency under the authority of the 1972 amendments to the Federal Water Pollution Control Act, specify the maximum monthly average effluent BOD and suspended solids requirements to be 30 and 45 mg/L, respectively. One should not overlook also the presence of refractory organics in the effluent that could eventually exert an oxygen demand on the receiving water. Tertiary treatment processes will be needed to polish the effluent to achieve the goal of zero pollution discharge.

1.2. Principles of Biological Oxidation

In the biological oxidation of wastewater, both synthesis and oxidation occur. Many groups of activated sludge microorganisms take part in carrying out the process. The important groups are described in Table 6.1.

In an activated sludge system that is properly operated, biological flocs are produced that incorporate most of the important groups of microorganisms. Metazoa and the fast moving protozoa are not part of the biological flocs because they can break away from them. Nevertheless, metazoa and protozoa constantly graze on biological flocs and are consequently found together in a highly stabilized wastewater. Although the ecosystem of an activated sludge process is complex, some general principles of biological oxidation can be applied (2–10).

1.2.1. Physical Adsorption

It has been observed that a fast initial removal of organics usually occurs when the wastewater is contacted with activated sludge in an aeration tank. This initial removal can be accomplished in a few minutes. The removal rate depends upon the wastewater characteristics and the volatile solids concentration of the activated sludge. The removal is interpreted as an adsorption phenomenon and was reported as early as 1939 by Ruchhoft (2).

The initial adsorption removes primarily discrete and colloidal particles. The adsorbed organic matter is subsequently oxidized or used in the synthesis of cellular components. It is important to recognize that, once adsorbed onto the biological flocs; the removal of organics from the wastewater is complete because the flocs are to be separated in the final clarifier. However, initial adsorption has little or no practical effect on dissolved organics. This explains why a contact stabilization process can take full advantage of initial adsorption only when

Table 6.1
Important groups of activated sludge microorganisms

Group	General characteristics	Identified organisms
Bacteria		
1. Zoolgea-forming	A growth from of various species, nonspore forming, motile, and capsulated rods, nonnitrifying, rapidly oxidize carbohydrates and produce ammonia from gelatin, casein and peptone, produce well-organized flocs	B. *subtilis*: aerobic spore former, g^+, nonnitrifying Alcaligenes: g^- rods, proteolytic Chromobacterium: g^- rods, proteolytic (Flavobacterium) Pseudomonas: g^- rods, prefer carbohydrates Achromobacterium: g^- rods Spirillum: spiral or curved rods not flocculating
2. Filamentous	Form loose flocs having poor settling characteristics	*Sphuerotilus natans:* (leptothrix, cladothrix) Bacillus mycoides Crenothrix Beggiatoa
3. Others	Specialized or controversial functional importance	Nitrifying bacteria: *Nitrosomonas, nitrobacter* Intestinal group: B. *coli*, *B. aerogenes*. some proteolytic effect and little action in carbohydrates
Fungi	Rarely abundant except under abnormal circumstances. filamentous	*Geotrichoides paludosus Pullularia pullulans Phoma Oospora (Geotrichum) Sporotrichum* Fusarium: non-filamentous. good floc forming
Protozoa	Feed on bacteria, but also use soluble organics at high concentrations	Ciliate: free swimming (paramecium), stalked (*Vorticella opercularia*) Sutoria: early ciliated, free-swimming stage and an adult stalked stage: feed on bacteria and protozoa
	Control bacteria population to effect better organic removal	Flagellata (Mastigophora): Flagellated, only found in poorly aerated environment
Metazoa	Higher forms of life often found in activated sludge; at times they may become so abundant as to be considered a factor in the ecological system. Feed on bacteria, protozoa, and algae	Rotifer: multicellular, rotating motion of two sets of cilia at one end. aerobic Nematode: aerobic worms can metabolize solid organics not readily degraded by other organisms

it is applied to the treatment of wastewater containing large amounts of colloids and easily adsorbed solids (11, 12).

1.2.2. Substrate Intake

With few exceptions, most organic compounds of high molecular weight (macromolecules such as polysaccharides and proteins) cannot penetrate into cells for use in cellular metabolism. Specific exoenzymes are excreted by the cells to digest the macromolecules adsorbed on the floc surface or in the wastewater. The digestive enzymes are hydrolases that catalyze the hydrolytic decomposition of their substrate. Examples of hydrolases are alpha-amylase (which breaks down glucose polymers), cellulase (for cellulose), proteolytic exoenzyme (breaks down peptide bonds in protein molecules) and lipase (which hydrolyzes fats and other esters).

Without digestive enzyme, only low molecular weight substances gain entrance into cells. There are indications that hydrophilic groups containing OH, COOH, and NH_2, or sulfonate, etc. with twelve carbons or under can pass through cell membranes, whereas hydrophobic groups with more than eight carbons cannot diffuse into cells. Nonhydrolyzed, hydrophobic compounds gain entrance in a different manner. According to McKinney (3) these hydrophobic compounds are attracted to the lipid fraction of the cytoplasmic membrane where they are soluble. By penetrating into the lipid fraction, the remainder of the molecule can be brought into the cell.

Not all diffusible substance can penetrate into cells. The cell can absorb and retain certain substances selectively while excluding or excreting others. Although all cells have this same general property, different organisms differ markedly from each other in their ability to accept certain specific organic nutrients from the wastewater for use in their metabolism. The selective nature of a cytoplasmic membrane is caused in part by its lipoprotein composition. There are specific "combining sites" in the membrane that effect the selective transport of particular compounds and ions into the cell. The possession of specific penetration or transport mechanisms plays an important role in the substrate specificity of many bacteria. Stainer (4) cites an example in which many bacteria are unable to oxidize citrates as organic nutrients simply because these compounds do not enter the cell. The same bacteria, however, possess all the enzymes necessary for citrate oxidation and produce citric acid constantly as an intermediate metabolite.

1.2.3. Intracellular Enzymatic Actions

Inside the cell, chemical transformations take place with the help of intracellular enzymes. The CoA portion of the enzymes react with the carboxyl group of short chain acids, amino acids or hydroxy acids, to form a CoA-acid complex. A number of reactions will follow. Most often, β-oxidation takes place in which enzymes remove H_2 and add H_2O to the organic molecules. The resulting acetyl-CoA will then enter the tricarboxylic acid cycle for terminal oxidation. Acetate metabolism is very common with microorganisms and it is the key intermediate for energy and synthesis according to McKinney (5). The tricarboxylic acid cycle, which is well accepted as the terminal oxidation scheme for acetate, is presented in

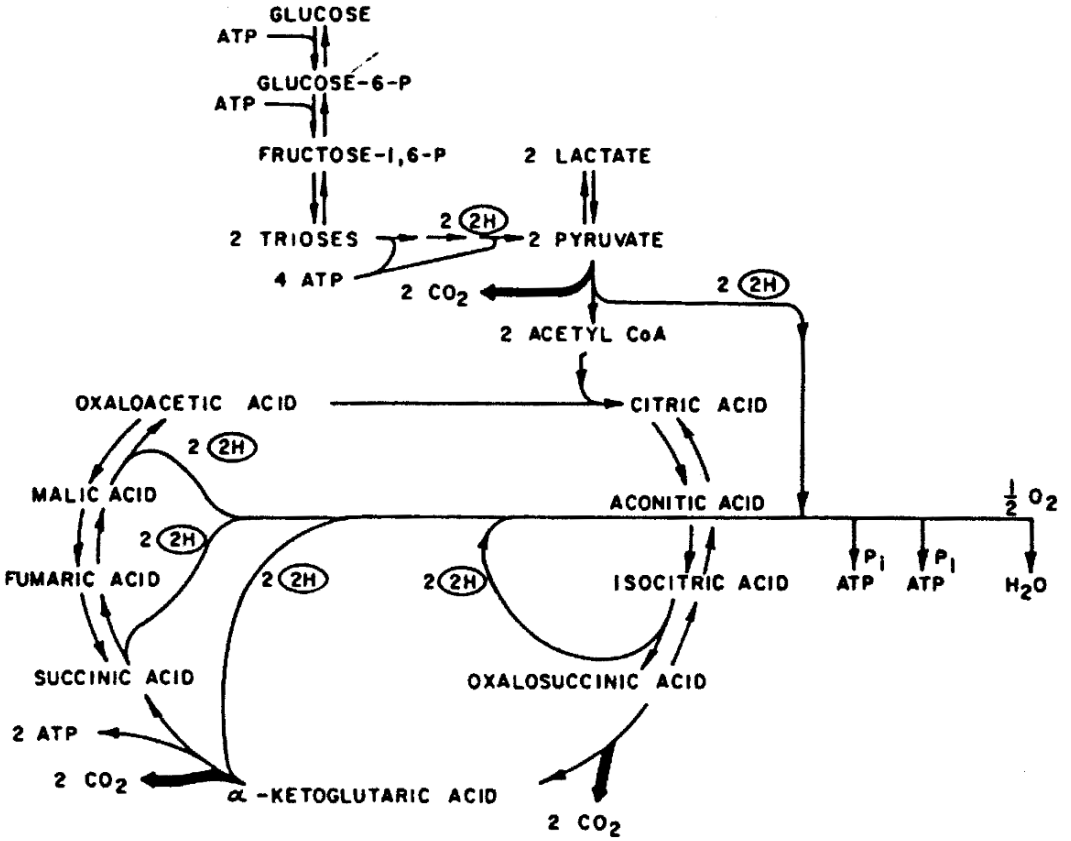


Fig. 6.1. Oxidation of glucose through acetate metabolism and trycarboxylic acid cycle.

Figure 6.1. Six CO_2 molecules are produced in three different steps to show the fate of the organic carbon in glucose.

With short chain alcohols, aldehydes, ketones, and amines, the reactions are directed towards conversion of the hydrophilic group to a carboxyl group so that reaction with CoA is possible. Chemical changes of organic substances in wastewater by microorganisms are summarized in Figure 6.2.

1.2.4. Hydrogen Transfer

The oxidation of organic compounds requires the removal of hydrogen. Activated sludge treatment, being an aerobic process, uses oxygen as the final hydrogen acceptor, as shown in Figure 6.1. Before this final step is taken, however, hydrogen removal is brought about by the coenzyme and cytochrome systems. The sequence of enzymatic reactions that mediate between the oxidation of a substrate and the reduction of oxygen is called the electron transport system.

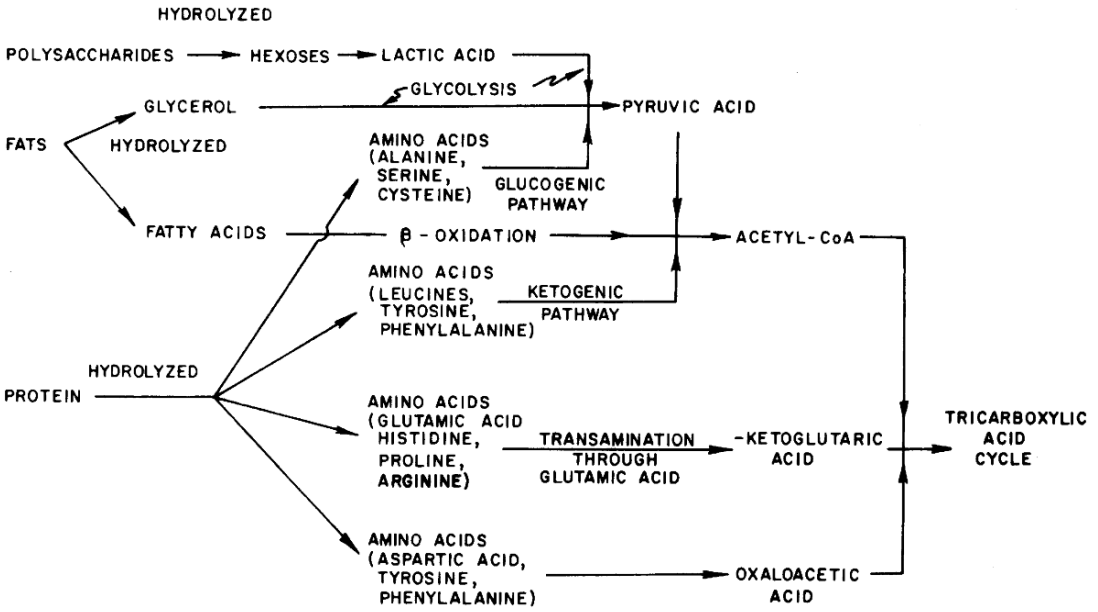
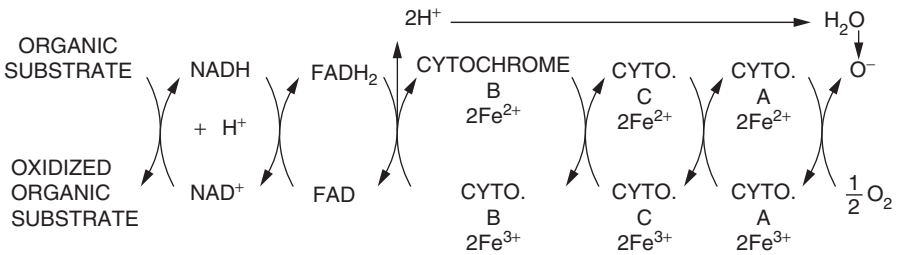


Fig. 6.2. General scheme of chemical changes by wastewater microorganisms.

For example, the oxidation of lactic acid by lactic dehydrogenase requires continuous regeneration (reoxidation) of NADH. NADH cannot be oxidized directly by molecular oxygen. Its regeneration requires a second coenzyme FAD:

$$\mathrm{NADH} + \mathrm{H^+} + \mathrm{FAD} \rightarrow \mathrm{NAD^+} + \mathrm{FADH_2}$$

where FAD is flavine adenine dinucleotide and NAD is nicotinamide adenine dinucleotide. FAD becomes reduced in the process and is regenerated by the reduction of a third catalyst, the third by a fourth, and so on until finally a catalyst that can react with oxygen becomes reduced. The general scheme of an electron transport system can be illustrated as follows:



The cytochrome complement associated with respiration in eukaryotic organisms is uniform, whereas many different cytochrome systems exist in bacteria. It should be noted that in the regeneration of FAD, there is a net liberation of hydrogen ions, because only electrons are

transferred to the cytochrome systems:

$$FADH_2 + 2Fe^{3+} \rightarrow FAD + 2H^+ + 2Fe^{2+}$$
$$4Fe^{2+} + O_2 \rightarrow 4Fe^{3+} + 2O^{2-}$$

The hydrogen ions are consumed in the subsequent reduction of oxygen to form water as a final product.

1.3. Energy Flow

Oxidation of organic substrates yields large amounts of energy. The electron transport systems provide for a smooth release of this energy. Some energy can be readily used by the microorganisms and the remaining can be converted into the phosphate bond energy of ATP for storage. The amount of energy available from oxidation of substrates depends on the nature of the substrate and on the metabolic pathways used by an organism. The potential energy available from the complete oxidation of glucose with molecular oxygen, according to McCarty (6), is 28.7 kcal/electron equivalent or 688,000 cal/mol of glucose. Conversion of glucose to alcohol and CO_2 has a potential energy of only 58,000 calories. The alcoholic fermentation of a molecule of glucose results in the net generation of two molecules of ATP. Assuming the bond energy of ATP is 7,000 cal, it can be calculated that approximately 25% of the potential energy can be tapped for use by the microorganisms. In the case of complete oxidation, one molecule of glucose yields 38 molecules of ATP or approximately 266,000 cal. The efficiency, at 40% of the potential energy, is high compared to mechanical systems.

The potential energy (Gibbs free energy change, ΔG) for a substrate oxidation reaction can be calculated from published data on free energy for half reactions (6). The following is a list of substrates commonly found in wastewater and their potential energy with complete oxidation in the presence of molecular oxygen:

Substrate	ΔG k cal/electron mol	Substrate	ΔG kcal/electron mole
Glucose	−28.7	Glutamate	−26.3
Fructose	−28.7	Butanol	−25.8
Lactose	−28.7	Benzoate	−25.6
Sucrose	−28.7	Butyrate	−25.5
Glycine	−27.1	Propionate	−25.3
Alanine	−26.3	Acetate	−25.3

In an activated sludge process, the substrates in the wastewater are assimilated by heterotrophic bacteria. Part of the energy is spent for supporting various metabolic activities. The bacterial population supports the growth of protozoa with a further energy loss as a result of the predator activities. A similar loss in energy takes place at the higher level (metazoa as predators). The useable energy in the process is therefore gradually diminishing. Given a long aeration time in the process, more energy is consumed and a higher effluent quality is obtained as a result.

1.3.1. Redox Potential and Electrode Potential

The chemical transformation brought about by biological treatment processes involves oxidation-reduction reactions. The general equation for oxidation-reduction potential is given as:

$$E = E_0 + (RT/nF)\ln[\text{oxidant}/\text{reductant}]$$

where E_0 is the potential when [oxidant] = [reductant], R is the universal gas constant, 1.99 cal/mol/deg, T is temperature in degrees Kelvin, n is number of electron moles transferred per mole of substrate used for energy, and F is Faraday, 23,060 cal/eV.

The electron activity defined as $pE = -\log[e-]$ is conceptually related to free energy and redox relationships. pE is a convenient measurement of the oxidizing intensity of a system at equilibrium and is related to the redox potential and Gibbs free energy according to Stumin (7):

$$pE = E(F/2.3RT) = -\Delta G/(2.3\ nRT)$$

or

$$pE = E/0.059 = -\Delta G/1{,}362\ n \quad \text{at } 25^\circ\text{C}$$

It can be seen that pE, as an intensity factor, is conceptually similar to pH. It represents the electron free energy level per mole of electrons. When the substrates in the wastewater are oxidized to release energy, aerobic system is characterized by a high pE value, indicating high energy availability whereas an anaerobic system is characterized by a low pE value indicating low energy availability.

For systems involving the coupling of electron transfer with hydrogen ion transfer, the general equation for oxidation-reduction potential becomes:

$$E = E_0 - 0.03\log[\text{oxidants}/\text{reductants}] - 0.03\ (\text{pH})$$

The above equation describes the redox potential for a thermodynamically equilibrated system. The activated sludge process is a steady-state system, but is thermodynamically not at equilibrium. Nevertheless, the pE concept and the general redox potential are useful in a qualitative manner to describe a biological treatment system. It is, however, suggested that the measured potential should be referred to as the "Electrode Potential, E_c" rather than "redox-potential" because the two have slightly different meanings.

A typical curve showing the relationship between E_c and substrate concentration in wastewater is shown in Figure 6.3. The curve shows that the positive E_c value increases with the degree of treatment. A well-stabilized wastewater effluent therefore should have a high positive E_c value. Dirasian (8) reported potential values ranging from +100 to +550 mV for aeration tanks of numerous activated sludge processes. The wide range of values should not surprise anyone because the measured potential depends on wastewater characteristics, temperature, pH, DO, efficiency of treatment, location where the reading is taken, and type of reference electrode used. If one measures the electrode potential over a lengthy period of time

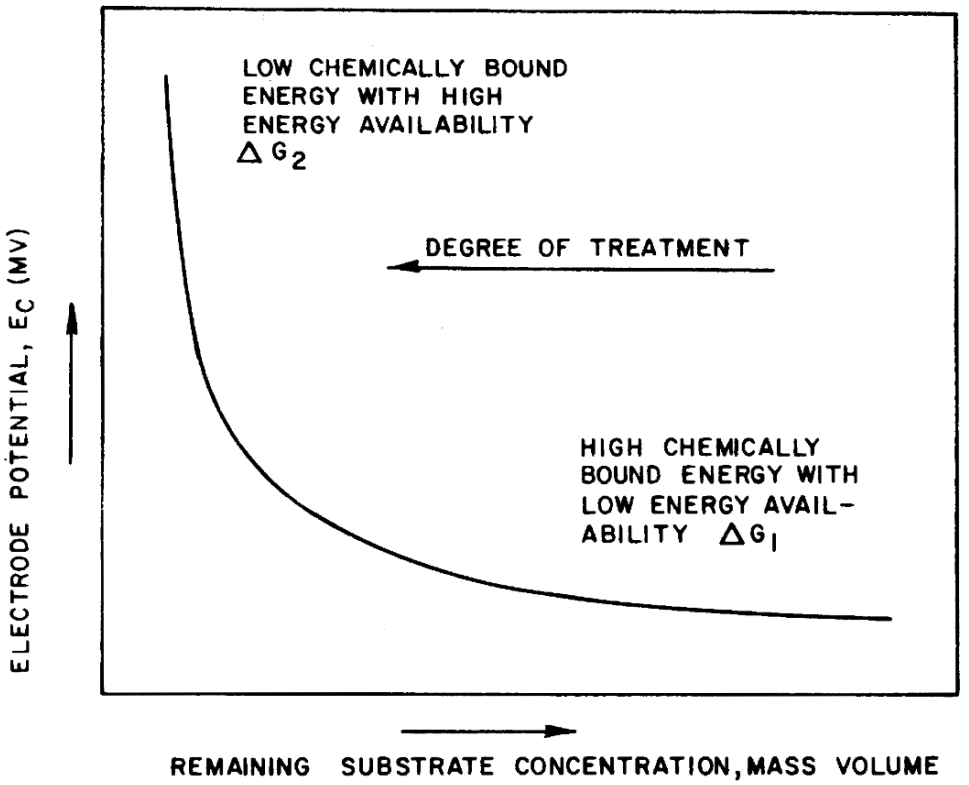


Fig. 6.3. Relationship between E_c and substrate concentration in wastewater.

in a treatment plant at a specific location, the range of values can be narrowed down and the potential should reflect closely the degree of treatment.

Measurement of E_c for monitoring the efficiency of treatment has many advantages over other monitoring techniques, (a) low cost, (b) real time monitoring and (c) no training required. In conjunction with DO and pH measurements, E_c monitoring can provide valuable information to aid the operation of activated sludge processes.

1.4. Synthesis and Respiration

In the process of biological oxidation of organic substrate, synthesis of cellular protoplasm takes place. The same enzymes that attack the substrate entering the cell also oxidize the cellular protoplasm continuously. Thus this latter action, often called endogenous respiration or self metabolism, occurs simultaneously with synthesis. When there is plenty of organic substrate available, synthesis exceeds endogenous loss, resulting in a net growth of cells. Conversely, when substrate is depleted, endogenous loss exceeds synthesis, resulting in net loss of cell mass in the system. The relative amount of synthesis and endogenous respiration depend mainly on the time of aeration. This is demonstrated in Figure 6.4. Increasing the aeration time will result in more endogenous mass loss. The effect of sludge age on the distribution

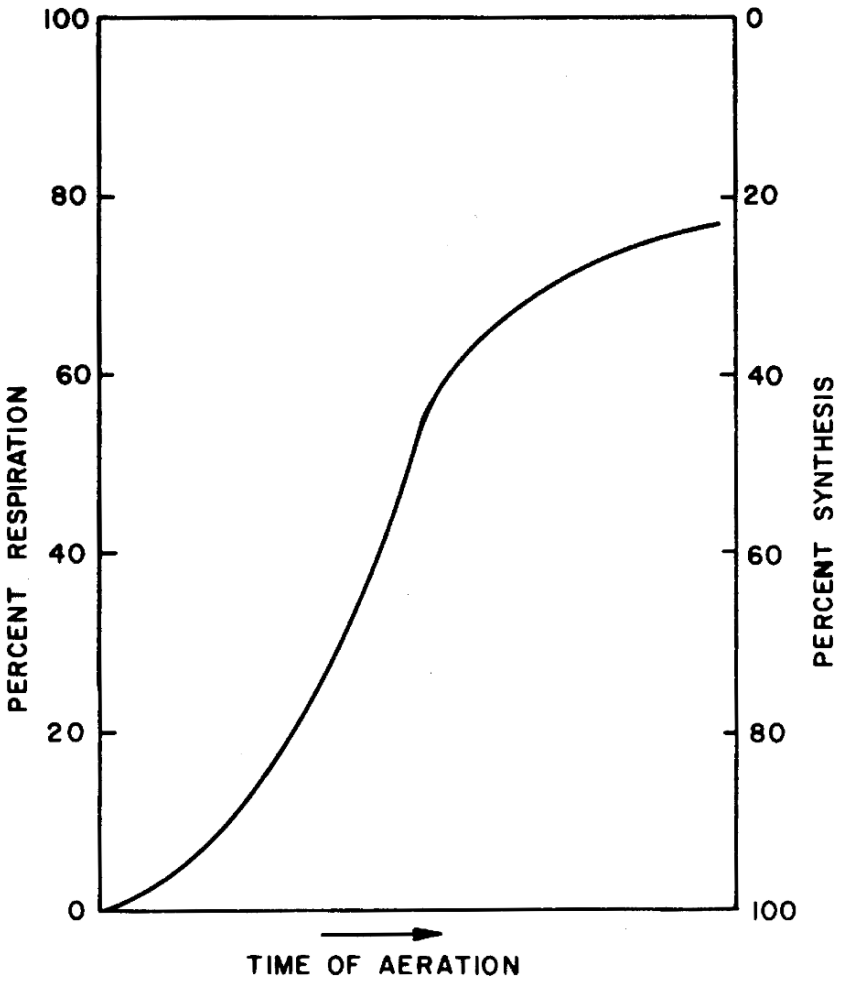


Fig. 6.4. Distribution of synthesis and respiration at varying aeration time.

of synthesis and respiration is more pronounced. Different activated sludge processes employ different aeration times and sludge ages. Consequently varying amounts of biological solids are produced. Factors affecting biological growth in the system also include the nature of substrate, the substrate in biomass ratio (F/M ratio), which is related to sludge age, substrate concentration and temperature. The growth of biological solids and its engineering control are discussed in later sections of this chapter.

2. SYSTEM VARIABLES AND CONTROL

The activated sludge process involves a highly complex biological system. To design and operate the treatment system properly, one needs to understand the system variables as well as the control alternatives that are available. Engineers have in the past identified the major

system variables that significantly affect the performance of an activated sludge process. Among these variables are: hydraulic retention time, organic loading, mean cell residence time (sludge retention time), aeration, mixing, and enzyme regulation. It is possible to control these variables, partially if not completely, through process design and operation. Discussions on these system variables, their relationships, and their significance in process control and performance are presented in this section.

2.1. Kinetics of Sludge Growth, and Substrate Removal

The success of an activated sludge process in producing a high quality effluent depends on the continuous growth of biological flocs having a good settling characteristic. The growth of these biological flocs is accompanied by organic substrate removal, where the rate of microbial growth and the rate of substrate use are interrelated. If one assumes that the Michaclis-Menten enzymatic kinetics can be applied to the substrate use by microorganisms in the process (13, 14), then

$$U = \frac{\mathrm{d}S/\mathrm{d}t}{X} = \frac{k_\mathrm{m} S}{K_\mathrm{s} + S} \tag{1}$$

in which U = specific substrate use rate, change of soluble substrate concentration per unit time per unit microbial mass; S = substrate concentration in mass per unit volume; X = microbial concentration in mass per unit volume; k_m = maximum rate of specific substrate use; and K_s = Michaelis-Menten constant, or half velocity coefficient, which numerically equals the substrate concentration when $U = 1/2\, k_\mathrm{m}$ in mass per unit volume.

Biological growth is the result of the coupled synthesis-endogenous respiration reactions described in the previous section. The net result can be expressed as

$$\mu = \frac{(\mathrm{d}X/\mathrm{d}t)}{X} = YU - b \tag{2}$$

in which μ = net specific growth rate, change of microbial concentration per unit time per unit microbial concentration, time -1, Y = growth yield coefficient, mass microbial growth per unit mass of substrate used; and b = endogenous or decay coefficient, time -1.

Considering a simple biological reactor with complete mixing and no sludge return, a microbial mass balance equation can be written for the reactor:

$$V(\mathrm{d}X/\mathrm{d}t) = V(YUX - bX) - QX \tag{3}$$

in which V reactor volume and Q = wastewater flow rate through the reactor in volume per unit time.

At steady state, i.e., $(\mathrm{d}X/\mathrm{d}t) = 0$, Equation (3) yields

$$D = 1/\theta = YU - b = \mu \tag{4}$$

which establishes the relationship between dilution rate (reciprocal of hydraulic retention time, θ) and the net rate of specific growth as well as the specific substrate use rate. It should be noted that $D = \mu$ at the steady state, because a constant microbial concentration X can be maintained in the reactor only when the net specific growth is continuously washed out.

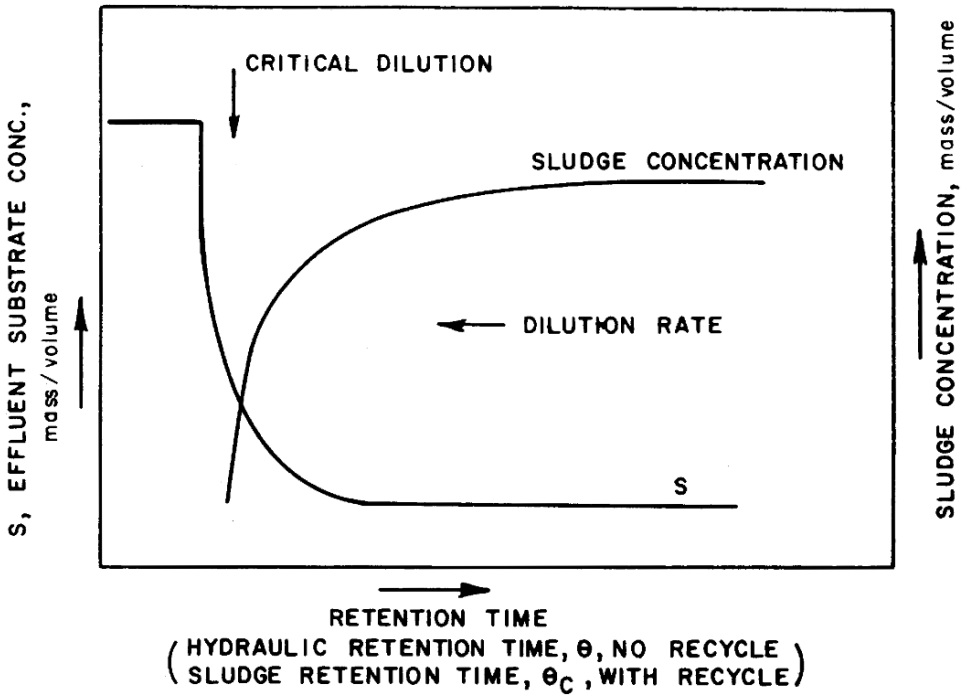


Fig. 6.5. Effect of hydraulic retention time on the performance of an activated sludge process (Complete-mix Model).

Engineers recognize the fact that to operate this system properly, the hydraulic flow should be controlled such that the dilution rate is smaller or equal to the net specific growth rate ($D \leq \mu$). When $D > \mu$, the microbial concentration decreases because of the high washout rate and system failure occurs. However, this problem can be minimized in a system with sludge return.

2.1.1. Complete-Mix and No Recycle Model

By combining Equations (1) and (4), the following equation is obtained:

$$S = \frac{K_s(1 + b\,\theta)}{\theta(Yk_m - b) - 1} \tag{5}$$

in which S = effluent substrate concentration at steady state. The effect of hydraulic retention time on the system performance can be illustrated by plotting S vs θ in Figure 6.5 which is similar to a diagram presented by Herbert (9).

From a substrate mass balance on the biological reactor, one can derive the following:

$$X = \frac{Y(S_0 - S)}{1 + b\theta} \tag{6}$$

in which X = effluent microbial concentration at steady state and S_0 = initial substrate concentration in wastewater. Assuming negligible amount of biological sludge leaving the final clarifier and knowing the rate of wastewater flow, the net sludge production rate can be easily calculated.

2.1.2. Complete-Mix and Sludge Recycle Model

With biological sludge recycled from the final clarifier, the mean cell residence time or sludge retention lime is longer than the hydraulic retention time. The sludge retention time is calculated as θ_c in the following:

$$\theta_c = \frac{VX}{Q_w\, X_r + (Q - Q_w)X_e} \tag{7}$$

in which Q_w = wasted sludge flow rate, volume per unit time; X_r = return sludge concentration, mass per unit volume; and X_e = sludge concentration in the treatment effluent from the final clarifier. By writing the mass balance equation for sludge in the entire system and assuming both X_e and X_o are in negligible amounts (X_o = sludge concentration in the primary effluent), one can develop the following:

$$\frac{1}{\theta_c} = \mu = YU - b \tag{8}$$

Following the same procedure in the development of working equations for the no-recycle model, one derives the following:

$$S = \frac{K_s(1 + b\theta_c)}{\theta_c(Yk_m - b) - 1} \tag{9}$$

and

$$X = \frac{\theta_c}{\theta}\frac{Y(S_0 - S)}{(1 + b\theta_c)} \tag{10}$$

One readily recognizes the similar nature of Equations (5) and (9). Equation (5) expresses the effect of the hydraulic retention time on system performance for a complete-mix no-recycle process as is shown in Figure 6.5. It is important to know from Equation (9) that the performance of a complete-mix with recycle system does not depend on hydraulic retention time. For a specific wastewater, a biological culture and a particular set of environmental conditions, all coefficients K_s, b, Y and k_m become constant. It is apparent from Equation (9) that the system performance is a function of θ_c. Thus it is possible to regulate θ_c to achieve good treatment efficiency without increasing the hydraulic retention time. This is basically the advantage of a recycle system over a no recycle system.

2.1.3. Plug Flow and Sludge Recycle Model

The plug flow model does not provide longitudinal mixing for adjacent elements of wastewater. The increasing microbial concentration and a concurrent decreasing substrate concentration along the axis of flow make the development of a kinetic model difficult. Lawrence (10) has developed a simplified model in which a constant microbial concentration,

X, in the reactor is assumed. This is believed to be valid as long as $(\theta_c/\theta) > 5$. Substituting the term X in Equation (1) with the average microbial concentration $\bar{X}$, integrating the equation over the hydraulic retention time of the wastewater in the aeration tank and simplifying, one obtains the following equation:

$$\frac{1}{\theta_c} = \frac{Yk_m(S_0 - S_1)}{(S_0 - S_1) + eK_s} - b \tag{11a}$$

in which $e = (1 + R)\ln[(RS_1 + S_0)/(1 + R)S_1]$, S_1 = effluent substrate concentration of the plug flow process. When R, the recirculated flow ratio Q_r/Q, approaches zero, $e = \ln(S_0/S_1)$, therefore

$$\frac{1}{\theta_c} = \frac{Yk_m(S_0 - S_1)}{(S_0 - S_1) + K_s \ln(S_0/S_1)} - b \tag{11b}$$

The equation applies as long as the volumetric recycle ratio (ratio of return sludge flow to influent wastewater flow) is less than unity. Equation (11b) shows that θ_c is a function of the influent as well as the effluent wastewater concentration, which is a unique characteristic of a plug flow process.

With a given set of values for the coefficients K_s, b, Y and k_m, one can calculate, from the equations given above, the sludge retention time θ_c required to produce a predetermined effluent substrate concentration. For the treatment of wastewater to obtain an effluent BOD of 20 mg/L or below, a shorter sludge retention time is required for the plug flow process. In other words, for a given θ_c value a plug flow process can obtain a lower effluent substrate concentration than that of a complete mix process.

2.1.4. Sludge Growth

Previously, it has been shown that the amount of cell synthesis depends on the length of time the cells are exposed to aeration. This phenomenon is expressed by Equation (4) for a no-sludge recycle process or Equation (8) for a sludge recycle process. Both equations state that a shorter retention time results in a higher specific growth rate and therefore more sludge growth. Because the effluent microbial concentration for a no-sludge recycle process is found to be $X = Y(S_0 - S)/(1 + b\theta)$ in Equation (6), the daily microbial sludge production can be calculated as:

$$X_\theta = \frac{VY(S_o - S)}{\theta(1 - b\,\theta)} \tag{12}$$

in which X_θ = daily microbial sludge production with no sludge recycle, mass/d when θ is expressed in days. A negligible amount of sludge loss with the secondary clarifier effluent is assumed in the calculation.

For a plug flow or complete mix process with sludge recycle, the daily sludge production is the amount of sludge wasted from the system, $Q_w X_r$. Neglecting the sludge in the secondary clarifier effluent, and combining with Equation (10), one finds:

$$X_{\theta c} = Q_w X_r = \frac{VX}{\theta_c} = \frac{VY(S_o - S)}{\theta(1 + b\,\theta_c)} \tag{13}$$

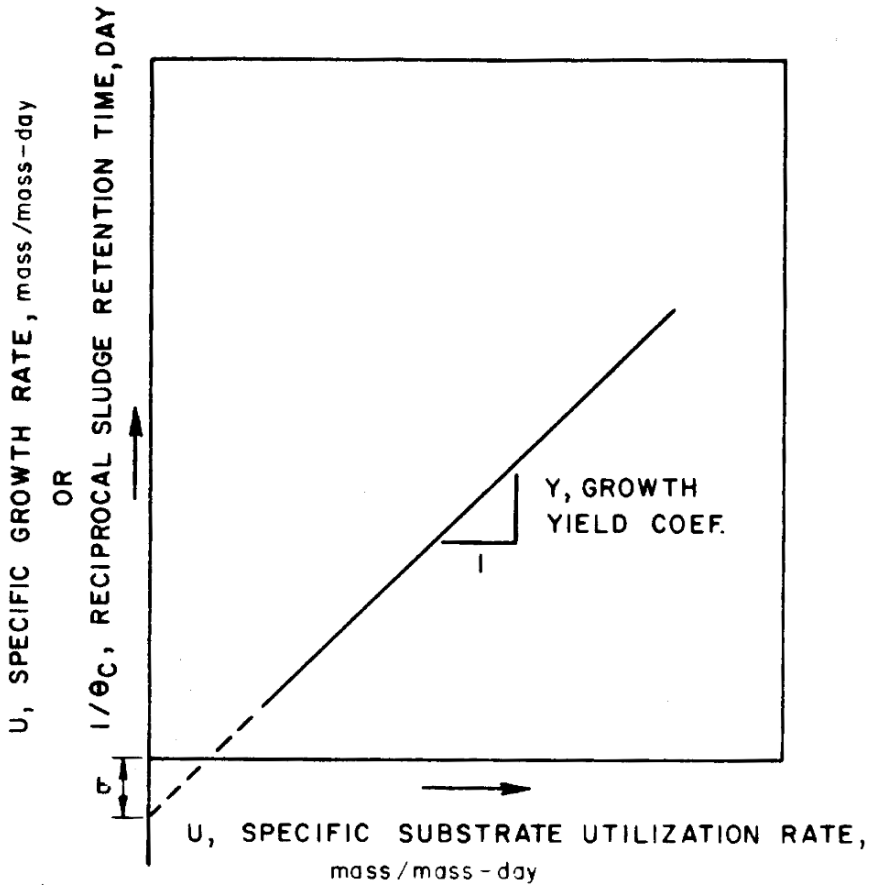


Fig. 6.6. Relationship between sludge retention, specific growth rate and substrate use rate.

The similarities between Equations (12) and (13) are apparent. For a plug flow system, the effluent substrate concentration S_1 should replace the steady-state substrate concentration S in the above equation.

2.2. *Process Variables, Interactions and their Significance in Process Operation and Performance*

The important process variables and their interactions have been presented mathematically. They are further delineated here to help engineers to see better their interactions. The significance of these variables in process operation and performance of the treatment system will be discussed in this section.

The single most important variable in activated sludge process is sludge retention time. The term sludge retention time is synonymous with sludge age and mean cell residence time. Equation (8) in the form of $1/\theta_c = \mu = YU - b$ relates sludge retention time to specific growth of biomass and specific substrate use rate. Figure 6.6 depicts their relationship. It is

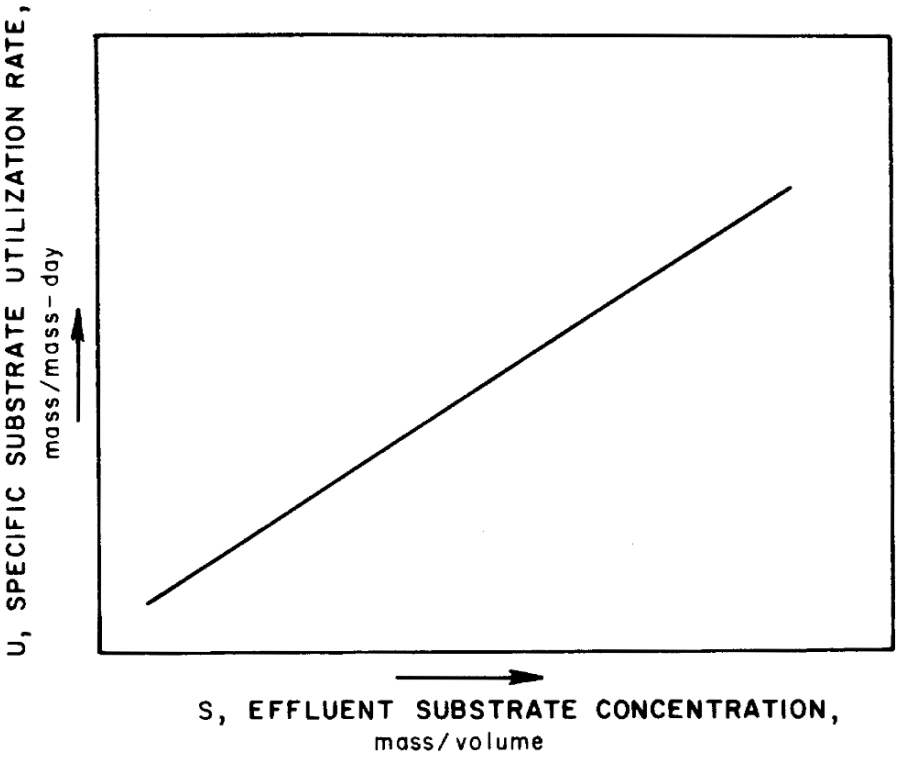


Fig. 6.7. Relationship between specific substrate use rate and effluent substrate concentration.

apparent from Figure 6.6 that within the limit of the capability of its biosynthesis, a young activated sludge (short sludge retention time) will grow faster and use the soluble substrate at a faster rate. This is desirable from the standpoint of minimizing the aeration time and aeration tank volume. Unfortunately the effluent substrate concentration increases with increasing specific substance use rate. Considering the rate of substrate removal as proportional to the existing biomass and substrate concentrations, $\mathrm{d}S/\mathrm{d}t = kXS$ where k is a proportionality constant, and writing a material balance around the aeration tank at the steady state,

$$0 = Q(S_\mathrm{o} - S) - VXkS$$

solving for S,

$$S = k\frac{S_\mathrm{o} - S}{XT} = kU \tag{14}$$

The relationship expressed by Equation (14) is illustrated by Figure 6.7.

To meet the requirement of a low effluent substrate concentration, therefore, a lower specific substrate use rate (or a longer sludge retention time) should be used. The terms food to biomass ratio (F/M) and specific substrate use rate have been used interchangeably by engineers. The two have slightly different meaning in definition in that F/M is the mass of substrate applied

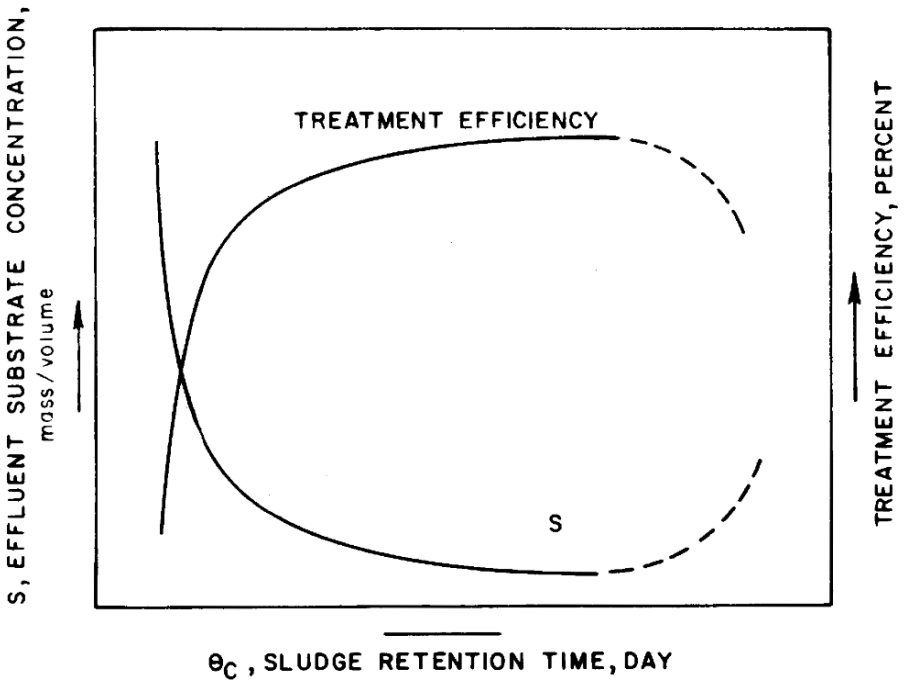


Fig. 6.8. Relationship between effluent substrate use and sludge retention time.

for unit biomass per day whereas U is the mass of substrate actually used by unit biomass per day. The value of U approximates F/M only when nearly all of the substrate applied to the system is used, a situation found in secondary treatment of organic waste in which a very high treatment efficiency is to be expected. However, it is not correct to assume such relationship when the system is not achieving high treatment efficiency.

At this point, it is necessary to reiterate the fact that all equations presented in the previous section and relationships between process variables aforementioned apply mainly to soluble substrates. With a significant portion of the substrate in the solid form of organics, the system may behave differently in its treatment performance. This fact is manifested in the contact stabilization operation in which the kinetics of growth and substrate removal are completely different from that of other modifications of activated sludge processes.

The effect of sludge retention time, θ_c on effluent substrate concentration can be found in Equation (9), as is illustrated in Figure 6.8.

The aforementioned kinetics also assumes a near complete removal of biomass from the final clarifier. In other words, the settling characteristics of the activated sludge is completely overlooked. Sludge retention time has significant effects on the sludge settling characteristics. A short sludge retention time and the resulting high substrate use rate will force the operation in exponential growth phase of the biomass. The biomass forms flocs poorly and does not settle well in the final clarifier resulting in low treatment efficiency. Conversely, with a prolonged sludge retention time and its resulting low specific substrate use rate the sludge is

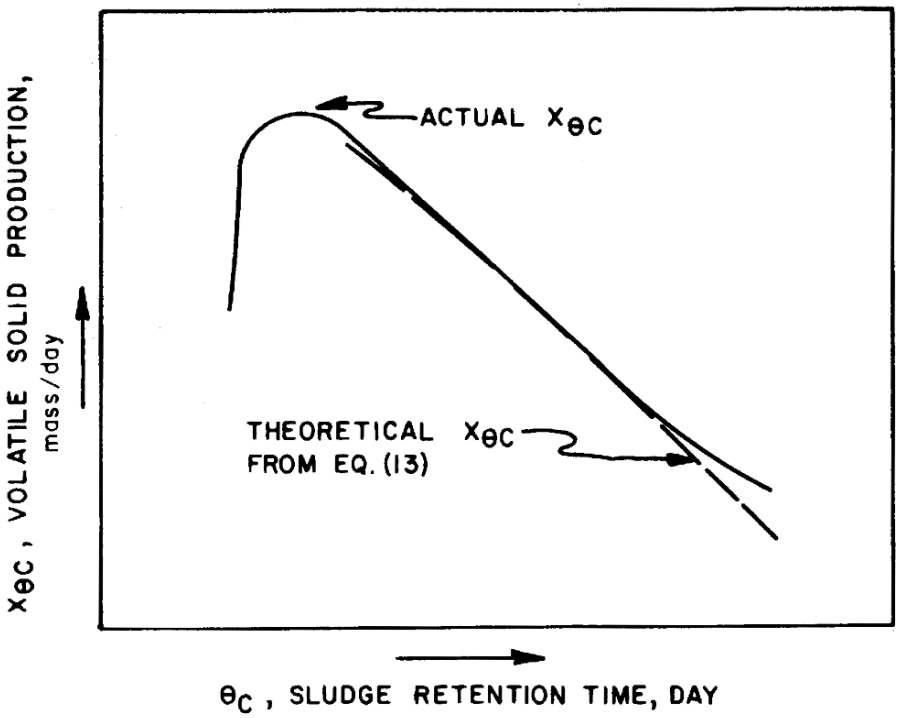


Fig. 6.9. Relationship between sludge production and sludge retention time.

subjected to an extended period of endogenous respiration and becomes inactive. As a result, pinpoint flocs will develop and settle poorly in the clarifier. In either case, the treated effluent contains a significant amount of BOD in the form of biological solids. Thus the treatment efficiency is low, as is illustrated in Figure 6.8. To produce the good settling flocs essential for a successful treatment, engineers select a θ_c value that gives neither a high nor low substrate use rate in the process operation. Suggested θ_c values for various activated sludge processes are presented later in this chapter.

Sludge retention time is normally controlled by wasting a portion of the settled sludge before returning to the aeration tank. To extend the sludge retention time, one needs to waste the biological sludge less often. This operation in effect generates less sludge from the treatment plant. This relationship between θ_c and sludge production is expressed in Equation (13) and is illustrated in Figure 6.9. It shows a significant drop of sludge production when the sludge retention time is very short. This is the result of poor sludge settling associated with a small θ_c value. The sludge leaves the system with the treated effluent in significant amounts and therefore reduces the amount that needs to be wasted to maintain a steady biomass in the aeration tank.

Recognizing the important process variables and their interactions in the operation, it is evident to engineers that they cannot obtain a high quality effluent and at the same time achieve fast substrate removal producing good settling flocs and a minimal amount of biological

sludge. One can optimize the performance and cost of the treatment for an existing process or can select among various modifications of the activated sludge process to suit special needs. The various process modifications will be discussed in Section 3.

2.3. *Aeration Requirements*

Air is supplied to the aeration tank to satisfy the biochemical oxygen demand in the process of organic oxidation. In addition, diffused air is required for turbulent mixing to keep the biological sludge in suspension and in intimate contact with the substrate. This is particularly true for diffused aeration, although mechanical aeration provides good mixing without relying on the diffused air in the wastewater. It is also believed that turbulent mixing by diffused air facilitates mass transfer of oxygen into the biological flocs and transfer of CO_2 and other waste products out of the flocs.

In the activated sludge process, the oxygen requirement consists of the amount of oxygen needed for both synthesis and respiration. Consequently one needs to know the ultimate BOD (BOD_L) of the wastewater, which can be obtained from BOD_5 using an appropriate conversion factor, The respiration oxygen demand is 1.42 g O_2/g VSS based on the empirical equation (15, 16)

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3$$

Because part of the VSS produced is wasted in the process operation for the control of sludge retention time, the respiration oxygen demand is reduced by an amount prepositional to the amount of wasted sludge. The theoretical oxygen requirement for an activated sludge process is therefore:

$$\text{Theoretical } O_2 \text{ requirement} = (BOD_L \text{ of wastewater used/d}) - 1.42\ (\text{VSS wasted/d}) \quad (15)$$

in which all terms are expressed in mass per day.

In practice, air is supplied to the aeration tank liquid to maintain a minimum dissolved oxygen concentration of 1 to 2 mg/L. The objective is to maintain a dissolved oxygen gradient across the liquid-floc interface to ensure an effective oxygen transfer into the biological flocs. The critical O_2 tension for the biological floc is believed to be in the neighborhood of 0.1 mg/L DO. Based on the geometry of floc particles, the passive transport of O_2 molecules through aggregates of living cells by diffusion has been mathematically resolved by Wuhrmann (17) as follows:

Spherical floc	$d^2 = (C_o - C_i)(24D/\alpha)$	(15)
Cylindrical floc	$d^2 = (C_o - C_i)(16D/\alpha)$	(16)
Plane floc	$d^2 = (C_o - C_i)(8D/\alpha)$	(17)
Parallel floc (biological film)	$d^2 = (C_o - C_i)(2D/\alpha)$	(18)

in which d = diameter of a sphere, cylinder or the thickness of a plane or parallel floc, cm; C_o = DO concentration at the surface of the floc (same as the DO concentration in the liquid medium), g/cm^3; C_i = DO concentration in the innermost cell of the floc, g/cm^3; D = diffusion coefficient, 5×10^{-6} cm^2/s at 15°C and α = specific consumption of O_2

molecules by the cells in the floc, g/cm^3/s and at 15°C; $\alpha_{min} = 1.0 \times 10^{-4}$ mg O_2/cm^3/s whereas $\alpha_{max} = 5 \times 10^{-3}$ mg O_2/cm^3/s.

With $C_i = 0.1$ mg/L. and a minimum $C_o = 1$ to 2 mg/L maintained throughout the aeration rank, biological flocs with maximum diameter of 340 to 490 μm can be adequately supplied with oxygen without interference of normal respiration. Because the average floc size in activated sludge process is approximately 200 μm or below, a DO concentration of 1 to 2 mg/L in the aeration tank is adequate.

Because the oxygen transfer efficiency in wastewater medium is very low, much more air is supplied to the aeration tank than is present in the liquid to satisfy the oxygen demand and to maintain the minimum DO of 1 to 2 mg/L. The transfer efficiency varies depending on the nature of the wastewater and the geometry of the aeration tank. Eckenfelder (18) reported transfer efficiency ranging from 6.4% to 24.8%. Most diffused aeration devices have transfer efficiency between 8% and 12%. For mechanical aeration devices, the average transfer is between 1 and 1.5 kg O_2/hp/h (2.2 to 3.3 lb O_2/hp/h). Detailed discussion on aeration and its design is presented in Chapter 4, Submerged Aeration.

Actual air requirements range from 3.75 to 15 m^3/m^3 of wastewater (0.5 to 2.0 ft^3/gal) depending on the strength of the wastewater. The volume of air required per unit mass of BOD removal therefore is a better basis for design. A range of 500 to 900 ft^3/lb of BOD removal (31 to 56 m^3/kg of BOD removal) has been recommended for the design of aeration in activated sludge treatment plants (19, 20). In addition, the ten states standards recommends a minimum air flow of approximately 3 cfm/ft of aeration tank length (17 m^3/h/m) for maintaining adequate mixing velocities to avoid deposition of solids.

2.4. Temperature Effect

The rates of all chemical reactions are affected by temperature. It is generally accepted that the rates of all biological reactions vary with temperature according to the van't Hoff-Arrhenius relationship. The integrated form of the van't Hoff-Arrhenius equation is:

$$\ln \frac{k_2}{k_1} = \frac{E_a}{R}\left(\frac{1}{T_2} - \frac{1}{T_1}\right) \tag{19}$$

in which k_2 and k_1 are reaction rates at temperature T_2 and T_1 respectively, E_a = activation energy in cal/mol, and R is the universal gas constant, 1.99 cal/mol (K). Most often a modified Arrhenius equation is used to predict the temperature effect (15, 21, 22):

$$K_T = K_{20}\theta_T^{(T-20)} \tag{20}$$

in which K_{20} is the cell growth rate or substrate use rate at 20°C and K_T is the corresponding rate at some temperature, T; θ_T is constant called be temperature coefficient. Values of θ_T for activated sludge process are generally in the range of 1.0 to 1.03.

Equation (20) is obviously oversimplified. The θ_T value is by no means a constant for all activated sludge processes. It varies even for a given activated sludge process depending on many factors. If one examines Equations (19) and (20), it can be seen that θ_T is a function of the activation energy. Inorganic chemical reactions generally have activation energies that are changed only slightly by temperature variations and their Arrhenius plots are essentially

linear. The effects of temperature on activation energy for biochemical reactions are far more complex. Activation energy may change in a system with a heterogeneous microbial population because of (1) a shift of predominant species as temperature changes, and (2) a change in nutrient substrates being used as the population shifts with changes in temperature (23). Other investigators have found that θ_T depends on substrate concentration, chemical nature of substrate, food to microorganism ratio, number of test temperatures used, method of chemical analysis and the procedure for evaluation of the rate constant (24–27). In addition the aeration process, as an energy controlled kinetic rate process with its rate depending upon temperature, has been found that the activation energy and therefore the θ_T value varies linearly with temperature (23).

It can be seen that the exponential form of temperature correction relationship, as is presented by the modified Arrhenius equation, is limited in its applicability because θ_T is not a constant, but varies with temperature and other factors. Over a moderate and narrow temperature range, the θ_T value does not change significantly. Wastewater temperature in activated sludge processes varies much less than in trickling filters and lagoons. A temperature variation of no more than $\pm 7°C$ from 20°C is expected for activated sludge processes in most treatment plants. Provided that this is true, Equation (20) can be used for temperature correction of treatment performance. It is suggested that a small θ_T value ($\theta_T = 1.0$) be used when the food to microorganism ratio is very low (0.2/d) and a high θ_T value ($\theta_T = 1.03$) be used when the *F/M* ratio is high (0.6/d or above).

3. SYSTEM MODIFICATIONS AND DESIGN CRITERIA

In the past, the conventional activated sludge process was used exclusively because engineers were handling relatively small volume and low strength wastewaters. Later in dealing with wastewater of greater volume and strength and with industrial wastewater in particular, which has a complex chemical nature, engineers were forced to operate the activated sludge process differently as well as to modify the process design to suit special needs. This led to the development of many process modifications that have become standardized today (28).

In this section, both the conventional activated sludge process and its modifications are discussed. The characteristics, removal efficiency and application and design parameters are listed for comparison (29).

3.1. Conventional Activated Sludge Process

The process employs a long rectangular aeration tank. Biological sludge is collected from the final clarifier. Part of the sludge is wasted from the sludge return line and the recycled portion, together with the wastewater influent (overflow from primary clarifier); enter the aeration tank at the head end. The mixed liquor is aerated for approximately 4 to 8 hours during which time synthesis and respiration take place with the organic matter being oxidized. The mixed liquor has a MLSS (mixed liquor suspended solids) of 1500 to 3,000 mg/L. The sludge settles in the final clarifier until there are 8,000 to 10,000 mg/L, when it is withdrawn off for partial return. The return rate is approximately 15% to 50% of the influent flow rate (Figure 6.10A).

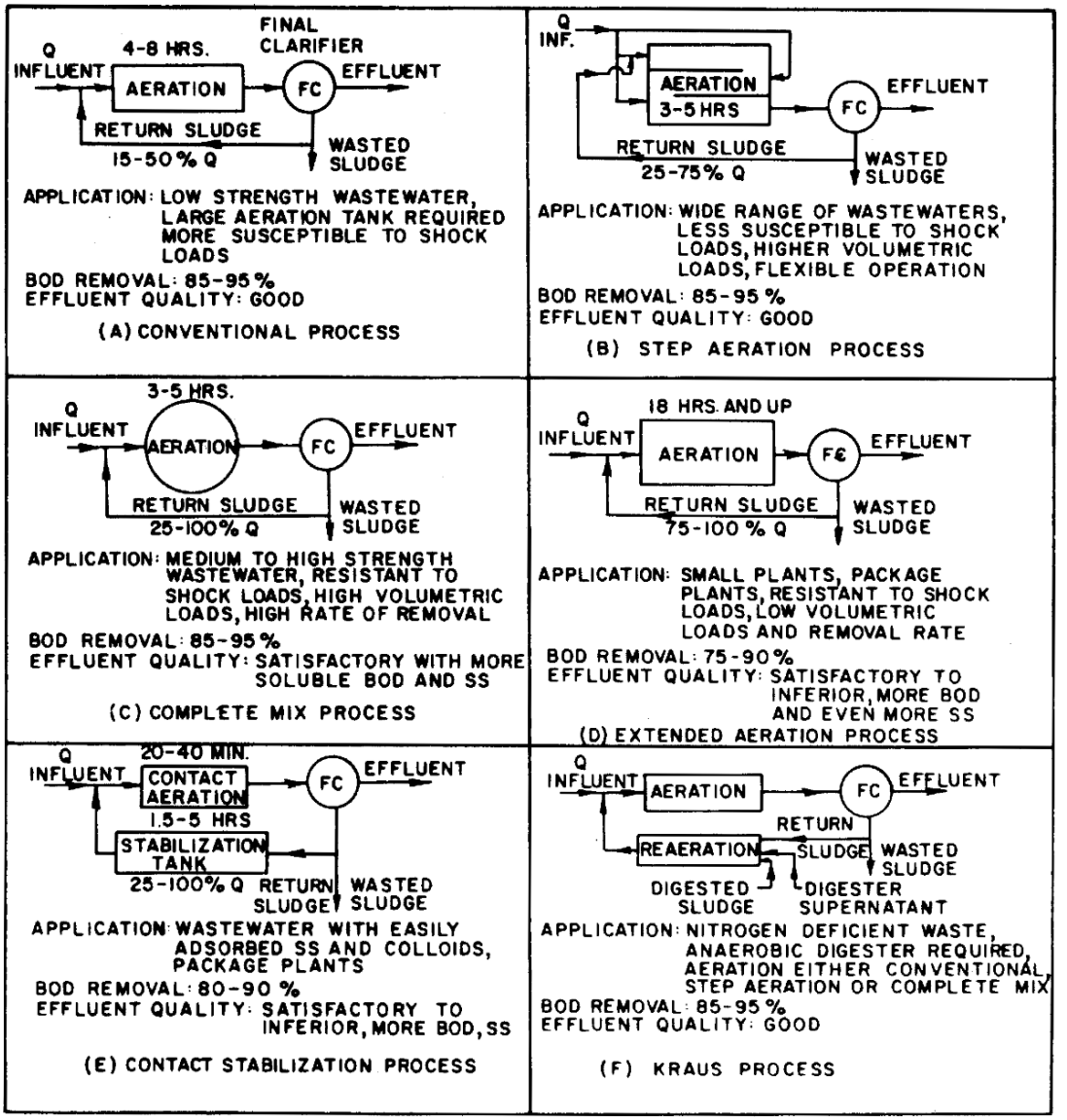


Fig. 6.10. Flow diagrams and applications of major activated sludge processes.

Hydraulically, the conventional process is operated in a quasi-plug flow mode using a long and deep rectangular tank. Some longitudinal dispersion and some short circuiting are expected and a true plug flow regime is not obtained. A true plug flow system is a more efficient process, as was discussed earlier. It is, however, more susceptible to shock loads if the aeration tank is divided into a series of complete mix reactors, Jenkins (30) has shown that improvement in treatment performance can be obtained without a major loss in the ability of

the system to handle shock loads. Toerber (31) reports that in response to a severe shock load, the plug flow operation in long rectangular tanks has a removal efficiency only 10% smaller than does the complete mix operation on a BOD basis. A conventional process in reality comes close to a series of complete mix reactors and consequently its treatment performance is one of the best among all activated sludge processes in use. Nevertheless, the conventional process is designed for and applicable to low strength wastewater only, for example, domestic wastewater. The requirement of a larger aeration tank, and consequently a larger land area, further inhibits its use.

3.2. *Step Aeration Process*

The step aeration process represents a significant improvement of the conventional process with very little physical modification. Return sludge enters the head end of the tank with a portion of the wastewater influent. The piping is so arranged that an increment of wastewater be discharged into the aeration tank at subsequent steps. In doing this, the waste load is more uniformly spread over the length of the aeration tank, resulting in better use of the oxygen supplied. More importantly, the load is more uniformly spread so that the system is much less susceptible to shock load. The biological sludge is maintained highly active throughout the tank so that more organics are removed in a shorter contact time. Consequently, a high waste load (volumetric loading, kg BOD/m^3 or lb BOD/1000 ft^3 of aeration tank volume) is possible, resulting in a significant reduction in unit size compared to the conventional process.

Either a long rectangular aeration tank or a rectangular tank subdivided into four or more parallel channels can be used. In either case, the first portion of the aeration tank can be used for reaeration of the return activated sludge alone. This adds to the flexibility of operation because recondition of return sludge by sludge aeration sometimes is essential for maintaining top treatment performance. The arrangement of subdividing the aeration tank into multiple parallel channels (Figure 6.10B) further adds flexibility to the process operation. In small cities and communities where fluctuations of wastewater flow and strength are expected, a portion of the aeration tank can be shut off to accommodate lower flow. This operation is very important if uniform treatment performance is to be maintained. Flexibility in operation provided is therefore an important feature of the step aeration process.

3.3. *Complete Mix Process*

The complete mix process is most useful for treatment of wastewater with fluctuating organic strength because it is least susceptible to shock loads among all activated sludge processes. For this reason, the process has found increasing popularity among those particular concerned with the treatment of industrial wastewater with moderate to high strength. A mechanically stirred reactor test simulates a complete mix condition (Figure 6.10C). When diffused aeration in a rectangular tank is used, the mixture of wastewater influent and the return sludge should enter the aeration tank at several points in a central location with the effluent going into channels on the sides of the aeration tank.

The process can be better described mathematically than other activated sludge processes (see Section 2). In general, both volumetric loading and F/M ratio are higher than that used in conventional and step aeration processes. A high MLVSS concentration is maintained to

accommodate the strong organic wastewater that leads to the higher sludge recycle ratio required, even though the sludge retention time is comparable to those of the step aeration and conventional processes.

It should be emphasized here that although comparable treatment efficiencies are obtained for the complete mix process, the conventional process, and the step aeration process, the higher *F/M* ratio employed in the operation of a complete mix process tends to leave more soluble in the effluent. Consequently an effluent with a lower quality is expected.

3.4. Extended Aeration Process

The extended aeration process is operated hydraulically, rather more like a complete mix than any other activated sludge process (Figure 6.10D). This is made possible by providing a very long hydraulic retention time, allowing the mixing of incoming fluids with those in the rest in the aeration tank with little or no chance of short circuiting. Wasting of sludge is done only periodically to avoid building up inert solids as well as to avoid excessive carryover of solids in the effluent.

The extended aeration time and the near complete return of all solids impose an endogenous growth condition on the system. Volumetric loading is very low and the sludge normally has an inferior settling characteristic. One can expect on the average a BOD removal efficiency of only 85% (32–34).

Because a long aeration time is employed, the process is generally applicable only to small treatment plants of less than 3785 m^3/d (1.0 MGD) capacity. Prefabricated package plants commercially available for housing subdivisions, small communities, institutions, schools, military bases, etc. generally use the expended aeration process. No primary sedimentation is provided and no exercise of return sludge control is attempted. The objective is to simplify the process in construction and in operation. Sludge production is very small because of prolonged endogenous oxidation, which minimizes the problem of sludge treatment and disposal. Needless to say, the oversimplified approach does not guarantee good treatment efficiency. In fact, more often than not, the process yields an effluent of lower quality. Most state water pollution control agencies are reluctant to approve the installation of such treatment plants, at least not as secondary treatment facilities.

3.5. Contact Stabilization Process

When activated sludge is contacted with wastewater in the aeration tank, there is an initial removal of organic matter that is often considered an adsorption phenomenon; this initial removal is rapid. The contact stabilization process takes advantage of the rapid adsorption characteristics of the activated sludge. The wastewater influent is allowed 20 to 40 minutes contact time with the return sludge in the aeration tank. During this period, much colloidal and suspended organic matter, and some dissolved organics are adsorbed by the sludge flocs to some extent. Although very little oxidation occurs in this brief period, the adsorbed organics are nonetheless removed as the sludge is separated from the treated effluent in the final clarifier. The settled sludge is then allowed 1.5 to 5 hours of aeration in a sludge stabilization tank in which the adsorbed organics are synthesized and oxidized. After the stabilization step, the sludge is partially returned to the incoming wastewater in the aeration tank (Figure 6.10E).

The advantage of the process is evident in that the shorter aeration time reduces the aeration tank volume considerably. Although the settled sludge needs aeration as well, the volume is small, whereas in the other processes the entire volume of the mixed liquor is aerated to achieve sludge stabilization. Aeration tank volume reduction by 50% over the conventional and step aeration processes is feasible. Many existing conventional plants are too small to accommodate the increasing volume and strength of the wastewater that they are converted to contact stabilization plants out of necessity. There are new treatment facilities, particularly package plants, designed and built as contact stabilization plants. The limit of the process application, for some reason, is not properly recognized by many engineers. The process should not be applied for the treatment of wastewaters whose organics occur mostly in the dissolved form. For municipal wastewater or even for domestic wastewater, laboratory tests should be performed to determine (a) the removable fraction of the organic matters and (b) the contact time and sludge concentration required to affect specified treatment efficiency.

Rich (35) assumes that adsorption removal per unit concentration of activated sludge is related to the concentration of the removable portion of the organics:

$$-\frac{\mathrm{d}S_\mathrm{a}}{\mathrm{d}X} = k_\mathrm{a} S_\mathrm{a} \tag{21a}$$

or

$$\frac{S_\mathrm{a}}{S_\mathrm{ao}} = -K_\mathrm{a} X \tag{21b}$$

in which S_a = concentration of remaining organics that can be removed by adsorption, mass per volume; S_ao = initial concentration of organics that can be removed by adsorption; X = initial concentration of activated sludge, and K_a = adsorption removal constant, which is equal to $0.434\,k_\mathrm{a}$ in Equation (21a).

The following procedure can be followed for a laboratory development of design criteria using Equation (21b):

1. Use four or more batch reactors, each with a different amount of acclimated activated sludge.
2. Allow a contact period of 20 min with the wastewater in each reactor.
3. Analyze the initial solid concentration, and the initial and final BODs of the wastewaters in each reactor.
4. Plot figures similar to Figures 6.11A and 6.11B to establish the removable fraction and removal constant at various solid concentrations.

The procedure may be repeated with different contact periods (30 minutes, 40 minutes ... etc.) to see whether a larger fraction of organics could be removed and whether such operation is economically justifiable.

From Figure 6.11, suppose a certain sludge concentration X is selected for operation of the adsorption process with a $(S_\mathrm{a}/S_\mathrm{ao})$ value of 0.1, then the total BOD removal of the process is $90\% - (90\%)(10\%) = 81\%$. If a total removal of 90% is desirable, it can be seen that 100% of the organics removable by adsorption is required, which is highly unlikely for any wastewater. The limitation of the process application is therefore self-evident.

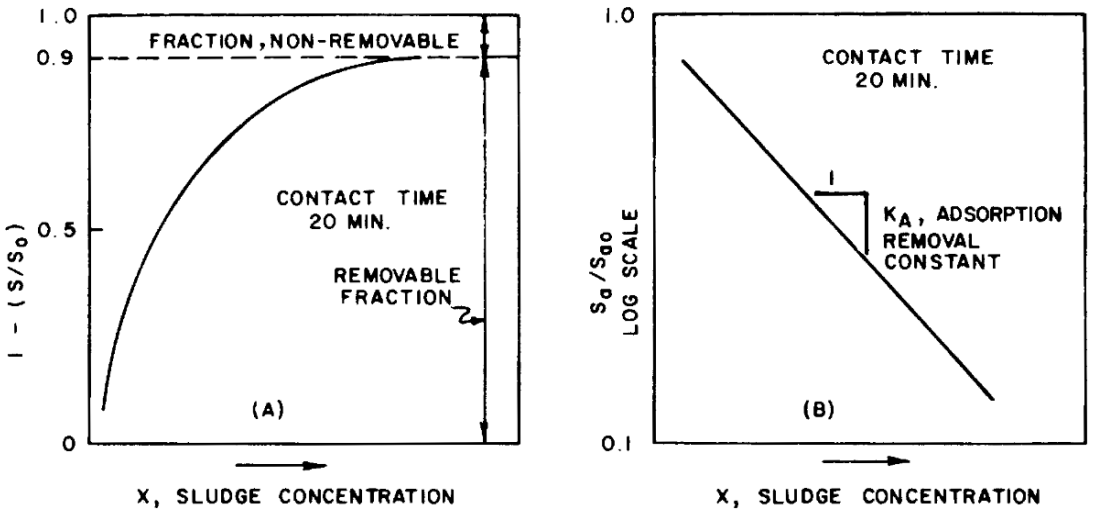


Fig. 6.11. Relationship between sludge concentration and removable BOD by adsorption.

3.6. Kraus Process

When nitrogen deficiency occurs in biological waste treatment and exogenous supply of nitrogen is expensive, the Kraus process can be useful in that an internal or endogenous supply of nitrogen is used to maintain the growth of active biomass. As shown in Figure 6.10F, some digested sludge and digester supernatant, together with a portion of the settled sludge from the final clarifier, are aerated in a reaeration tank. The released ammonia nitrogen from the sludge and the supernatant is converted to nitrate. When the content from the reaeration tank is introduced into the aeration tank with the wastewater, nitrate serves as the supplemental nitrogen source for synthesis. There is an improvement of the settleability of the MLSS also because of the presence of the inert fraction of the aerated, digested sludge. Additions of nitrogen in the form of ammonia, nitrate, or urea can be added to the reaeration tank for better control of nitrogen supply if needed.

3.7. Design Criteria

From past experience in the activated sludge process operation, design parameters have been established for the various processes. Table 6.2 lists the significant design parameters for various activated sludge processes (36). Design parameters for modified aeration and high-rate aeration from this source are omitted here for reasons to be given later.

For comparison purposes, Tables 6.3 and 6.4 are presented to show the current design criteria for activated sludge processes in New York and Illinois States respectively. It can be seen that whereas the design parameters of the two states tend to be slightly conservative, they are in general in conformity with those suggested in Table 6.2.

Table 6.2
Design parameters for major activated sludge processes

Process modification	*F/M* ratio, kg BOD/kg ML VSS/d	Parameter			Sludge retention time d	Hydraulic retention time h
		Volumetric lb BOD/10^3 ft^3/d	Loading, kg BOD/m^3/d	MLSS mg/L		
Conventional	0.2–0.4	20–40	0.32–0.64	1500–3000	5–15	4–8
Step aeration	0.2–0.4	50–60	0.64–0.96	2000–3500	5–15	3–5
Complete mix	0.2–0.6	50–120	0.80–1.90	3000–6000	5–15	3–5
Extend aeration	0.05–0.15	10–25	0.16–0.40	3000–6000	20–30	18–36
Contact stabilization	0.2–0.6	60–75	0.96–1.20	1000–3000	5–15	0.5–1.0
Kraus process	0.3–0.8	40–100	0.64–1.60	2000–3000	5–15	4–8
Pure oxygen system	0.25–1.0	100–250	1.60–4.0	6000–8000	8–20	1–3

Table 6.3
Design criteria for activated sludge processes for New York state – aeration tank capacities and permissible loadings

Process	Plant design flow MGD	Aeration period h (based on, design flow)	Plant design lb BOD_5/d	Aerator loading, lb. BOD_5/1000 ft^3/d	MLSS[a], lb/BOD_5, lb/d
Conventional	To 0.5	7.5	To 1000	30	2/1–4/1
	0.5–1.5	7.5–6.0	1000–3000	30–40	2/1–4/1
	1.5 and up	6.0	3000 and up	40	2/1–4/1
Modified or "high rate"	All	2.5 and up	2000 and up	100	1/1 (or less)
Step aeration	0.5–1.5	7.5–5.0	1000–3000	30–50	2/1–5/1
	1.5 and up	5.0	3000 and up	50	2/1–5/1
Contact stabilization,	To 0.5	3.0 (in contact zone)[b]	To 1000	30	2/1–5/1
	0.5–1.5	3.0 to 2.0 (in contact zone)[b]	1000–3000	30–50	2/1–5/1
	1.5 and up	1.5 to 2.0 (in contact zone)[b]	3000 and up	50	2/1–5/1
Extended aeration	All	24	All	12.5	as low as 10/1 to as high as 20/1

[a] MLSS/BOD_5: normally recommended values at the ratio of mixed liquor suspended solids under aeration (lb) to BOD loading (lb/d).
[b]Contact zone: 30% to 35% of total aeration capacity. The reaeration zone comprises the balance of the aeration capacity.

Table 6.4
Design criteria for activated sludge processes for the state of Illinois – tank capacity and permissible loadings

Process type	BOD and SS removal %	Design flow MGD	BOD loading lb/1000 ft^3/d	Detection time h	MLSS/BOD lb/lb/d	Detection time h	Surface settling rate gpd/ft^2	Weir overflow rate gpd/ft
Extended aeration	70	All	To 20	24	10/1 to 20/1	(To 0.05 MGD) 4.0	300	5000
						(0.05–0.15 MGD) 3.6	300	5000
						(0.15 MGD up) 3.0	600	5000
Contact Stabilization	90	To 0.5	30	3.0 mix, 6.0 reaer.	2/1 to 5/1	3.6	500	5000
		0.5–1.5	30–50	2.0–3.0 mix, 4.0–6.0 reaer.		3.0	600	10,000
		1.5 up	50	1.5–2.0 mix, 3.0–4.0 reaer.		2.5	700	10,000
Step aeration	90	0.5–1.5	30–50	7.5–5.0	2/1 to 5/1	(To 0.05 MGD) 3.0	600	5000
	90	1.5 and up	50	5.0		(0.5–1.5 MGD) 2.5	700	10,000
						(1.5 MGD up) 2.0	800	10,000
Complete mix	90	All	35	6.0	2/1	2.0	500	5000
Conventional	90	To 0.5	30	7.5	2/1 to 4/1	3.0	600	5000
		0.5–1.5	30–40	7.5–6.0		2.5	700	10,000
		1.5 and up	40	6.0		2.0	800	10,000

3.8. Other Processes

3.8.1. Tapered Aeration

Tapered aeration is only a slight modification of the conventional process in that diffusers are spaced close together at the head end of the aeration tank so that the higher oxygen demand is met by the more abundant oxygen supply. The spacing of diffusers increases toward the tank outlet as the oxygen demand is lowered rapidly. It is believed that the process can eliminate oversupply of oxygen at the outlet end and thereby inhibit the growth of nitrifying organism that further tax the oxygen supply. The belief that less air is required in tapered aeration with consequent lower operating costs is unfounded because the reduction of air used at the outlet end is cancelled out by the increase in air used at the head end. However, a better BOD removal performance can be expected in a tapered aeration process because of its more effective use of the air.

3.8.2. Modified Aeration

This process uses shorter aeration times, down to 1.5 to 3 hours. A low MLSS concentration is maintained at approximately 500 mg/L, resulting in a high F/M ratio. The purpose is to operate the system at the exponential growth phase of the activated sludge for fast removal of BOD. The short aeration time allows only 60% to 75% BOD removals. Furthermore, the sludge in the exponential growth phase does not flocculate properly. The net result of the poor settling characteristics of the sludge is a high effluent suspended solids concentration. Because of this inferior performance, the process is primarily suited to intermediate treatment and should not be designed as a secondary treatment unit. In fact the modified aeration process was originally planned as a modification of the old activated sludge treatment plants to accept wastewater with the flow rate and strength above their design capacities. The effluent quality is far below the secondary effluent standards specified by US EPA today.

3.8.3. High Rate Aeration

Similar to the modified aeration process, a high rate aeration process uses short aeration times, 0.5 to 2 hours, to take advantage of the fast BOD removal at the exponential growth phase of biological sludge. A high MLSS concentration (4000 to 8000 mg/L) is maintained to increase the sludge retention time in the system. Most often the system is operated with a very high volumetric loading (up to 500 lb/1000 ft^3) and a very high substrate use rate that poor settling of sludge in the final clarifier occurs. Even though a BOD removal of 75% to 85% can be obtained, the effluent still contains a significant amount of soluble and solid BOD that the effluent quality is no better or inferior to that of the modified aeration.

3.8.4. Step Feed

In all *st*andard process modifications that use plug flow mode operation, the return sludge enters the head end of the aeration tank. Balmér et al. (37) developed a process in which the preaerated return sludge is returned to the aeration tank in equal portions at various locations. In the standard step aeration process, part of the wastewater entering the outlet end of the aeration tank will have a short retention time compared to the part entering the outlet end of the aeration tank. The BOD removal will not be uniform, and a lower efficiency should be

expected at the outlet end. With step feeding of sludge, more biomass exists at the outlet end to ensure efficient BOD removal. Compared with conventional and step aeration processes, the step feed process maintains a higher MLSS concentration with a better adsorption property, and thus can receive a higher BOD loading.

Balmér's experience showed good quality effluent at high *F/M* or volumetric loading. An 85% BOD removal was obtained at 4 kg BOD/m^3/d (250 lb BOD/1000 ft^3/d) loading and 66% to 79% removal at 9.6 kg BOD/m^3/d or 600 lb BOD/1000 ft^3/d loading. The process is able receive 10 times as much volumetric loading as an extended aeration process with equivalent treatment performance.

3.8.5. Oxidation Ditch

Using a shallow channel with a circular path, the wastewater is introduced into the ditch continuously or intermittently for a long hydraulic retention time. The screened wastewater is aerated by an aeration rotor and circulates at about 30 to 60 cm/s (1 to 2 ft/s). The process is operated in a similar fasion to an extended aeration process or an aerated lagoon with equivalent performance (see Chapter 12).

4. COMPUTER AID IN PROCESS DESIGN AND OPERATION

The state-of-the art for predicting performance and for the design of activated sludge processes needs much improvement. Most mathematically derived formulas have unnecessary constraints for simplification. Lack of consideration of population dynamics and their effect on sludge settling ability as well as on the overall performance of the process is obvious. Most designs and operations have traditionally relied on formulas and procedures derived from experience. Nevertheless, engineers equipped with the formulas describing the relationships of the important parameters in the activated sludge process (Section 2) have developed computer programs to simulate the process to assist in process design and operation. Through these computer programming techniques, engineers may develop better understanding of the process, and our ultimate goals of better control and optimization of treatment performance can be achieved.

4.1. Prediction of Performance

Smith and Eilers (38) have developed two digital computer programs, one of which is steady-state model of the conventional activated sludge process (CSSAS). The CSSAS is flexible enough to simulate any of the model modifications. A hydraulic flow diagram for the program is shown in Figure 6.12. By dividing the aeration tank volume into multiple subaerators of equal volume in series, the hydraulic mixing regime for any process modification can be found somewhere between the plug flow and complete mix conditions by assuming that the aerator behaves like some number of completely mixed tanks in series. The CSSAS program provides for setting the number of tanks in series to any number between one (complete mix) and ten ($n = 10$ for plug flow). The stabilization tank is included for simulating a contact stabilization process. It is omitted from the diagram if another process modification is used.

The CSSAS program is a computational scheme programmed to solve mass balance and rate equations describing the activated sludge process. The program allows simulation of

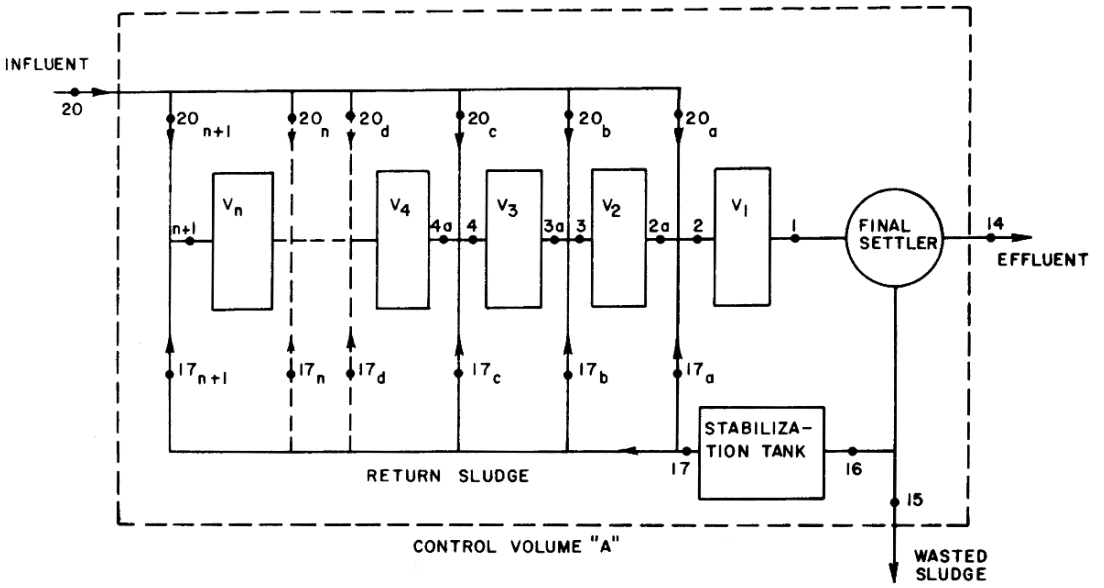


Fig. 6.12. Flow diagram for generalized computer model for steady-state performance of the activated sludge process. (*Source*: US EPA ref. (38)).

conventional, high-rate, extended aeration, step aeration and contact stabilization modifications. The computational flow diagram for the program is shown in Figure 6.13.

In the computation, a basic growth equation (Monod equation) similar to that of Michaelis-Menten Equation (1) is used. Inputs to the program include: activated sludge characteristics k_m, K_s, Y and b; characteristics of the influent stream, including $Q(20)$, $S(1,20)$, $S(2,20)$, *SBOD*(20), *VIS*(20), total aerator volume (VAER); the number of equal volume subaerators (NTKS); compaction ratio for the final clarifier (URSS) and effluent VSS (SSOUT).

Using the symbols in Smith's work (38) to rewrite Equation (1)

$$U = \frac{S_{n+1} - S_n}{X_n DT} = \frac{1}{Y}\frac{K_r S_n}{K_s + S_n}$$

Rearranging the equation, one obtains the following:

$$S_{n+1} = S_n + \frac{K_r S_n X_n DT}{Y(K_s + S_n)} \tag{22}$$

Similarly rewriting Equation (2)

$$\mu = K_n - K_e = \frac{X_n - X_{n+1}}{X_n DT}$$

Rearranging the equation, one gets the following:

$$X_{n+1} = X_n + K_e X_n DT - \frac{K_r S_n X_n DT}{K_s + S_n} \tag{23}$$

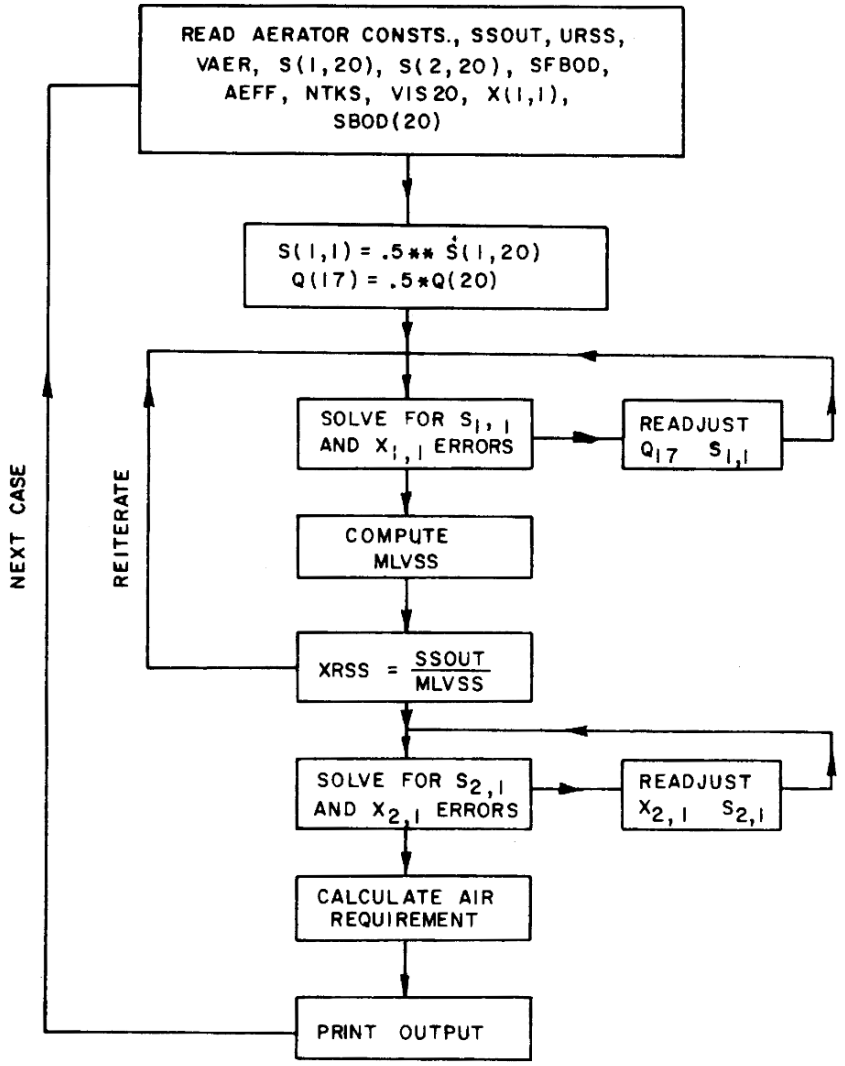


Fig. 6.13. Computational flow diagram for CSSAS activated sludge program. (*Source*: US EPA ref. (38)).

in which $DT = V/(Q_{20} + Q_{17})$. The mass balance relationships expressed by Equations (22) and (23) are used to compute concentrations of 5-d BOD and biomass at stations (1) through (NTKS + 1); (n) is the station number downstream of each subaerator. The Bolzano or "halfing" technique is used for all iterations as shown in the CSSAS flow diagram (Figure 6.13). Some computed results using the CSSAS program are reported by Smith in a graph form which is replotted here as Figure 6.14. It is interesting to note that a plug flow aerator performance better than that of complete mix aerators is confirmed in this plot.

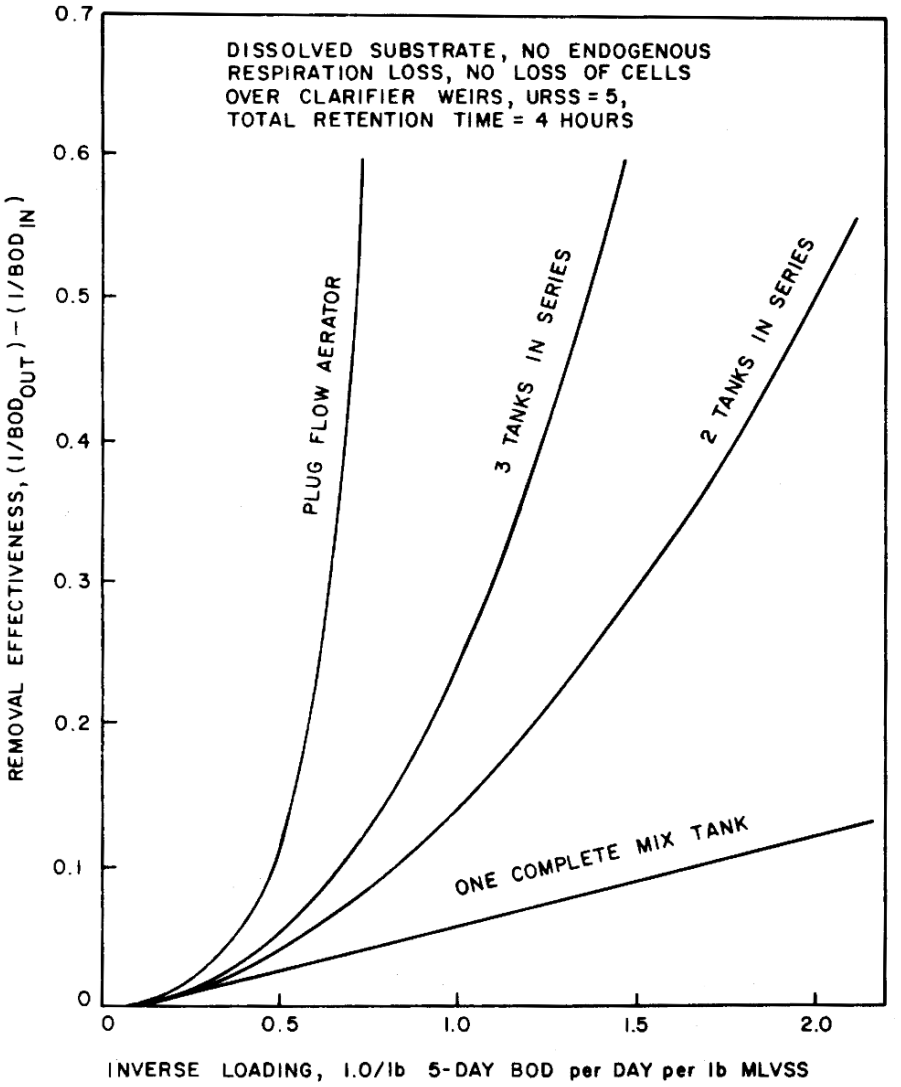


Fig. 6.14. Activated sludge process performance with plug flow aerator and 1, 2 and 3 tanks in series. (*Source*: US EPA ref. (38)).

4.2. Computer Program for Process Design

For activated sludge process design, Eilers and Smith (39) have developed, based on the performance model described earlier (38) and other work (40), a digital computer program for preliminary design of wastewater treatment systems. This computer program (FORTRAN; IBM 1130) can be used to compute the quasi-steady state performance and cost of groups of treatment unit processes arranged in any practical configuration. Each subroutine computes

the performance and cost of a single unit process, such as the activated sludge aeration tank and its final settler (AERFS).

Input to the (AERFS) subroutine includes, among others, demand effluent BOD, MLSS in the aerator, water temperature, BOD rate constant for sizing the aerator, DO, compaction ratio (URSS), design overflow rate for the final settler, and so on. As output, volume of the aerator, MLVSS, the ratio of solids concentration of effluent to MLSS in the aerator, the allowable load of influent BOD, the surface area of the final settler, the sludge return ratio for the aerator, the air requirement and the volume of the return sludge stream, among others, are included.

In the cost subroutine, the construction cost index, the amortization rate and period, the labor cost, the land and power costs and so on are fed into the program. The computation will yield total capital cost, total amortization cost, total operation and maintenance cost, total treatment cost, and other pertinent information.

The objective of the aforementioned computer programs is not to increase the accuracy and reliability of prediction and design. The programs are only as good as the few basic equations [Equations (22), (23), and others] with many constraints. However the influent stream vector and decision variables as inputs to the program can be changed within wide ranges and results can be quickly obtained. Consequently, an engineer can optimize the performance and treatment cost with little effort.

4.3. Computer Aid in Process Operation

Engineers have paid more attention to the design than to the operation of the activated sludge process. The process is a versatile and highly flexible one that demands carefully controlled operation. Unfortunately most operators do not have a proper understanding of the process and it has thus gained a reputation of being difficult to operate.

An activated sludge system can be divided into two distinguishable subsystems, the aeration tank and the final clarifier. The performance of the biooxidation process in the aeration tank and the resulting biological sludge characteristics affect the settleability of sludge in the final clarifier. On the other hand the performance of the final clarifier and the resulting return sludge characteristics affect the performance of the aeration tank. An attempt to incorporate sludge settling properties to define the relationship between sludge retention time θ_c and sludge volume index SVI was made by Bisogni (41). A lack of adequate consideration of pinpoint floc formation and of the poor settling associated with larger θ_c values is obvious in the study. Furthermore, SVI is considered today a poor parameter for the characterization of the performance of a sedimentation tank.

Recognizing the limitations of the mathematical models based on simple growth equations, a more practical approach has been taken to model operation formulas. Lacroix and Bloodgood (42) first developed mathematical models with linear polynomial approximation based on the available operation data of many activated sludge treatment plants. From these equations, they then developed a computer program to assist plant operation in controlling and optimizing treatment plant performance (43). Table 6.5 taken from their work (42) shows the predicted MLSS, return sludge SS, effluent SS, effluent BOD, and SVI based on equations expressed in terms of BOD loading, sludge age (sludge retention time), amount of wasted sludge, influent flow, wastewater temperature, and so on, based on the operational experience

Table 6.5
Predictive models and correlation coefficients for fort wayne indiana activated sludge plant[a]

Models	Number of observations	No. of terms in equation		Equations	Correlation coefficients	
		Initial	Final		R	R^2
SS in aeration tank effluent	47	27	7	$\text{WTSO} = 262.313 + 0.38385\ (\text{WTSR}) - 1.45935\ (\text{T}) + 0.000470\ (\text{WTSR})^2 + 0.38918\ (\text{WTSI})^2 - 0.01536\ (\text{WTSR})(\text{WTSI}) + 0.98277\ (\text{WTSI})(\text{SA}) - 29.24715\ (\text{WTSI})(\text{RT})$	0.96	0.92
Return sludge SS from sedimentation tank	47	20	6	$\text{YR} = 4{,}939.66 - 1{,}079.94375\ (\text{SA}) + 5.12657 + (\text{SOR}) + 1.62401\ (T)^2 - 6{,}278.26312\ (\text{R} + \text{W})^2 - 0.03450\ (X)(T) + 0.22006\ (X)(\text{SA}) + 3.30725\ (\text{X})(\text{R} + \text{W}) - 115.11847\ (\text{T})(\text{R} + \text{W}) + 439.31533\ (\text{SA})(\text{R} + \text{W})$	0.94	0.88
SS in plant effluent	47	20	6	$\text{WTSE} = 1.993 + 0.000019653\ (\text{SOR})^2 + 0.04415 + (\text{SA})^2 + 0.00347\ (T)^2 - 0.0000375\ (\text{WTSR} + \text{WTSW})(T) - 0.0003867\ (\text{SOR})(\text{T}) - 0.01672\ (\text{SA})(T)$	0.75	0.56
BOD in plant effluent	47	27	5	$\text{WTBE} = -1.69875 + 1.56600\ (\text{WTSE}) + 0.22166\ (\text{WTBI}) - 0.085932\ (\text{WTSE})^2 - 0.049725\ (\text{WTSE})(\text{WTBI}) - 0.00116844\ (\text{WTSI})(T)$	0.88	0.77
SVI	47	27	4	$\text{SVI} = -947.01 + 6461.08630\ (\text{RT}) - 8{,}218.69134\ (\text{RT})^2 - 0.000630\ (X)(T) + 4.50163\ (T)(\text{BL})$	0.76	0.58

[a] For definition of terms and notation, see Nomenclature/Definitions of terms.

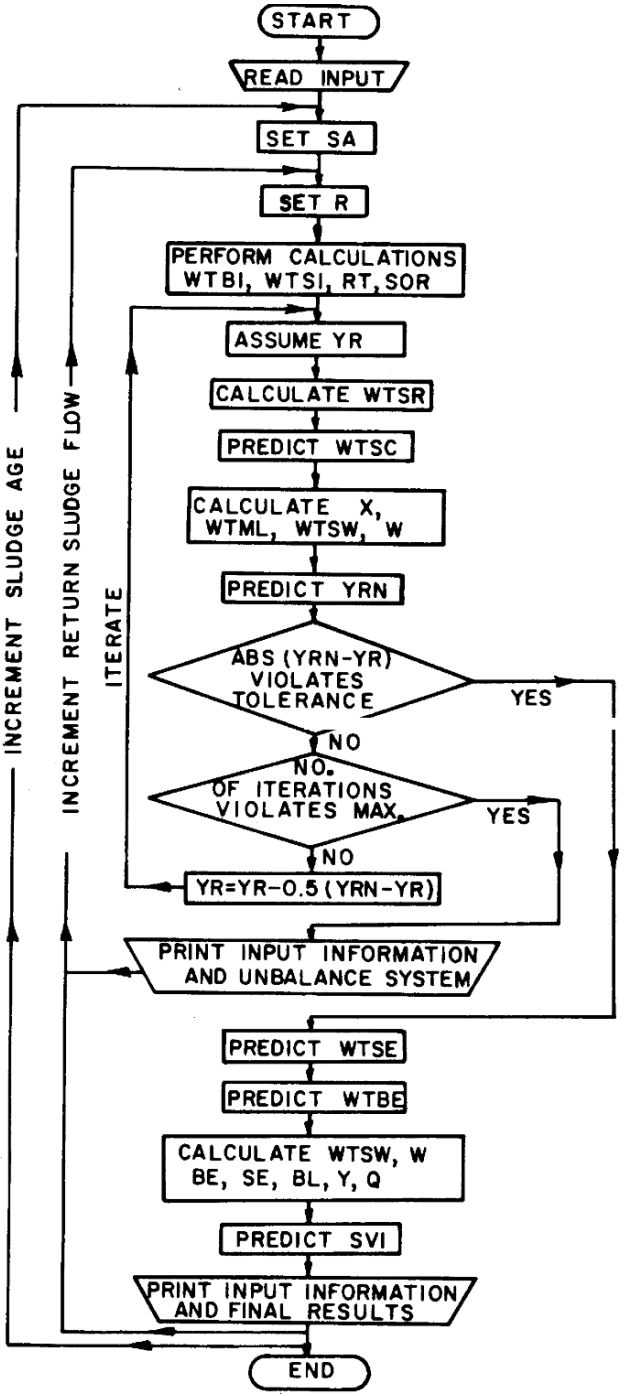


Fig. 6.15. CASSO flow diagram (43) (For definitions of terms and notations, see NOMENCLATURE).

at a particular treatment plant. Sensitivity analyses help to find the relative importance of many operational parameters on system response. Correlation coefficients in the table show that the effluent BOD can be predicted fairly accurately. Predictions of SVI and the closely associated effluent SS are not as reliable.

After identifying sludge age as the single most important parameter in process operation, Lacroix and Bloodgood (43) developed the CASSO program with its flow diagram shown in Figure 6.15. In addition to influent flow, BOD, SS, temperature, volume of aerator, and surface area of final clarifier, the sludge age is read as input to the program. Assuming the return sludge concentration later, both wasted sludge quantity and return sludge concentration can be computed using predictive equations. If the predicted return sludge concentration does not match the originally assumed value, correction based on half of the difference of the two values will be applied for reiteration until the difference is lower than or equal to a predetermined level. Other predictive equations will then be used to compute effluent SS, BOD, and SVI.

One of the uses of this CASSO program is to help the operator to select the return sludge flow or wasted sludge flow that will result in a predetermined treatment performance when the influent stream vectors are known. Because low effluent BOD and SS, small SVI, and minimal production of sludge cannot be achieved at the same time, another use of the program is to optimize the performance. In a long run, a substantial amount of cost saving can be realized. The success of this predictive control model depends on rapid and accurate assessment of the influent characteristics and equally important depends on the accuracy of the predictive equations. The availability of a long recorded history of operational data is a prerequisite with this approach.

5. PRACTICE AND PROBLEMS IN PROCESS CONTROL

Successful operation of activated sludge processes has long been recognized as difficult to achieve. Poor operational practice defeats a good design. In this section, current operational practice and the problems associated with it are discussed. Some possible remedies are suggested.

5.1. Wasting Sludge, Feedback and Feed Forward Control

Most often, the activated sludge treatment plant operator attempts to maintain a constant mass of activated sludge in the aeration tank with which to treat the incoming wastewater. In doing so, the operator needs only to monitor the average MLSS concentration in the aeration tank. For an influent stream with a relatively steady flow rate and strength, this control is adequate. A treatment plant with very long hydraulic retention time and complete mixing to equalize the fluctuation of the influent stream (i.e an extended aeration process) can use this simple control. With a fluctuating influent stream, a constant mass of sludge in the aeration tank yields a changing F/M ratio. Consequently, the specific substrate use rate and the resulting sludge retention time θ_c also change. The treatment efficiency will change accordingly. In some treatment plants, the uncontrollable diurnal variation in influent stream

flow rate and BOD may be several hundred percent of the mean values. A significant change of F/M ratio and θ_c could easily lead to process failure.

Another control method is to return the sludge at a rate proportional to the influent wastewater flow. In general, the strength of municipal wastewater is in proportion to its flow rate. It can be easily seen that when a high flow and stronger wastewater enters the treatment plant, the higher organic loading needs a proportional increase in added biological sludge to maintain a steady F/M ratio as well as a steady θ_c value. Recycling the settled sludge at a rate proportional to influent flow therefore is an effective control. The method is most simple in that no sludge concentration determination is required. This method of control is limited in its application to municipal wastewater treatment only.

In controlling both F/M ratio and θ_c the best approach is to monitor influent and effluent BOD, influent and effluent SS, MLVSS, return sludge concentration as well as influent, and return sludge flow rates. This allows the operator to see whether the FIM ratio is within the limits of design values (see Table 6.2) and if the θ_c value [calculation based on $\theta_c = (V + V_s)X/Q_wX_r$, see Example 1] falls within the limits of design values (Table 6.2). The successful use of this control method requires a great deal of laboratory work. Usually either one of the chemical oxygen demand test (COD), total oxygen demand (TOD), or total organic carbon test (TOC) is used in substitution for the BOD test because of the delayed action of the BOD test, making it unacceptable for control purposes.

Another alternative control method is the hydraulic control first proposed by Garrett (44). The operator returns all the settled sludge from the final clarifier while part of the aeration tank mixed liquor is taken out continuously to a separate clarifier from which the settled sludge is wasted entirely. Because the wasted liquor has the same sludge concentration as in the aeration tank, the sludge retention time θ_c is simply $(V + V_s)X/Q_w$. This allows the operator to estimate the Q_w for any desired θ_c value in the system. There is no need for determination of the BOD and the return sludge concentration. For a complete mix system, the waste liquor can be taken anywhere from the aeration tank. Burcheett et al. (45) have suggested a special waste-activated sludge manifold to be installed for quasi-plug flow system such as conventional and step aeration processes. It is necessary to bear in mind that an idealized performance of the final clarifier is assumed in this method of control. In other words, settling of the biological sludge should be complete so that all of it can be returned for θ_c control. If settling is poor because of a surge of slow, sludge bulking, or any other reason, with a resulting loss of solids in the effluent, the sludge retention time will gradually decrease and treatment efficiency will be lower.

In most cases, operators exercise the control of sludge retention time and F/M ratio by varying the amount of wasted sludge in response to decreasing treatment performance. A process control in response to output signal is called feedback control. Unfortunately the biological response of the process to rapid fluctuations in input conditions is very slow. A feedback control is an ineffective control process. On the other hand, feed forward control responds to input variations, which is more sensitive in process control. Westberg (45), Brett et al. (46), and Davis et al. (47) showed that a dramatic improvement in the dynamic performance is feasible by using feed forward control. Davis et al. (47), using a digital computer simulation method, showed that the best feed forward controller for a completely mixed

process is proportional for flow and derivative for concentration changes. The simulation does not include the dynamic performance of the final clarifier, however. Klei et al. (48) revealed in their pilot plant study that when the pilot reactor is operated at yield coefficient and hydraulic retention time characteristic of municipal plants, the treatment performance could be held steady when return sludge is controlled as a function of the strength of the influent stream. However, at low yield coefficients and shorter hydraulic retention times, the result is unsatisfactory. The dynamics of the settler become controlling, and the concentration of recycle solids is not adequate to improve performance.

With a properly designed activated sludge process, the best process control yet is to regulate the influent stream so that the flow rate and organic loading are within the design limits. This in effect is to prevent adverse operating condition leading to poor performance. An analog computer simulation study by Ott (49) shows that although return sludge (with a considerable reserved sludge storage capacity) could control the process, it is more effective to control by reducing the fluctuation of influent BOD and even more effective to control by reducing the fluctuation of influent flow rate. A complete mix influent diversion and storage tank can modulate both the influent flow rate and influent BOD. This is similar to the use of two complete mix aerators in series with the first unit being used to modulate the flow rate and BOD to the second unit. Future design of activated sludge processes should consider such an approach.

5.2. Bulking of Sludge and Rising of Sludge

Bulking is a phenomenon of overgrowth of certain undesirable microorganisms that do not settle readily in the final clarifier. Filamentous microorganisms are normally found in bulking sludge. These include, among others, Crenothrix. Phragmidiothrix, Clonothrix, Leptothrix, Sphaerotilus, Toxothrix, Thioploca, Thiothrix, Leucothrix, Pelonema, Beggiatoa, Vitreoscilla, Microscilla, Bactoscilla, Peloploca, Nocardia, Streptomyces, Micromonospora, and Bacillus (50–52). Also Geotrichum was reported by Pipes (53) and Schofield (54), and Zoophagus was found by Cooke (55). In addition, Flavobacter, Flexibacter, and Haliscomenobacter were found by Rensink (56). Bulking of sludge leads to loss of sludge organisms in the effluent, which may eventually lead to a process failure (57).

Overgrowth of filamentous forms of microorganisms occurs under various circumstances. In general, most of the filamentous organisms cited previously exist in municipal sewage in minute quantities. Nonfilamentous forms are most abundant and predominate in the activated sludge system because they compete better. When wastewater characteristics change, resulting in low pH, low temperature, high carbohydrate concentration, nitrogen deficiency, high chloride content and the presence of some species of heavy metal, of sulfur, or of petroleum products, the environment of growth may be found to favor some of the filamentous organism. Consequently, they grow in abundance and bulking of sludge occurs. Improper aeration, resulting in low oxygen content and/or high F/M ratio, may also lead to bulking of sludge. Although the causes are many, it is very difficult to identify one or two major ones for any particular plant that has sludge bulking problems. Jones (58) and Poon (58, 59) reported that such an occurrence is usually caused by a combination of many factors.

Correction of sludge bulking can be made by using either one or a combination of the following methods:

1. ***Identifying the Organism(s) Causing the Problem***
   *Farquhar's work (51) and Cooke's laboratory gu*ide (60) provide valuable information on isolation and a systematic key to the identification of these organisms. Because the growth conditions and requirements are known, corrective measures can be taken to inhibit their growth. The work, however, is tedious and requires considerable training. Most treatment plants do not have the personnel and equipment to carry this out.
2. ***Investigating abnormalities in influent stream characteristics and operation***
   Examples of these include low oxygen content in the aeration tank (a minimum of 1.5 to 2.0 mg/L DO should be maintained anywhere in the tank), low pH, nitrogen deficiency (much lower than that to maintain a 20/1 BOD removal to available N ratio), high *F/M* ratio, septic return sludge, heavy metals, shock loads of high chloride content, and so on. Appropriate steps should then be taken to eliminate the possible causes.
3. ***Application of chemical treatment as a remedy***
   Although polyelectrolytes are effective in precipitating the bulking sludge, the high cost prohibits their use. Chlorination of return sludge is effective and inexpensive. The chlorine dosage for control may vary from plant to plant depending on the degree of bulking, causative organisms, return sludge concentration, and so on. Wells (61) reported that a dosage of 0.3% to 0.7% by weight (dry weight of return sludge) stoped sludge bulking at the San Antonio Treatment Plant. The danger of over chlorination should not be overlooked, because it may kill off most other organisms, causing a process failure. At the aforementioned dosage, equivalent to 15 to 20 mg/L of chlorine, the contact period (a few minutes to 5 hours in the return sludge aeration tank) is not critical. Chemical treatment with hydrogen peroxide to control sphaerotilus overgrowth also has been reported to be successful by Caropreso (62). Depending on the SVI value in the bulking stage, dosages of 40 to 200 mg/L H_2O_2 are required over several days to correct the bulking problem completely. During the period when chemical treatment is applied, the effluent may be turbid. The effluent will be clear again once the bulking problem is completely corrected.

Rising of sludge occurs when settled sludge in the final clarifier floats to the surface after a brief settling period. When nitrification is completed in the aeration tank, both nitrite and nitrate are formed. Denitrification will occur at the bottom of the clarifier, where nitrites and nitrates are converted to nitrogen gas anaerobically. The nitrogen gas will carry the sludge mass to the surface because of the added buoyancy of the attached gas bubbles (63). A method of control is to reduce aeration so that nitrification does not occur. Nitrification is not required in an activated sludge treatment plant when the process is used as a secondary treatment. Another control method is to withdraw the settled sludge faster so that the chance of denitrification is greatly reduced.

6. CAPITAL AND OPERATING COST

In predicting the future cost of building and operating wastewater treatment plants, engineers rely on cost indexes. When prices of building materials are stable relative to the national economy and free from fluctuations caused by local conditions, the method is reliable. Construction cost indexes do take on a new importance as inflation soars. In the following

section, cost estimates based on the standard indexes for activated sludge treatment plants are given. A sample worksheet is included to illustrate the proper procedure to be used. Methods of increasing the accuracy of cost estimates will be discussed.

6.1. Traditional Cost Estimates

Various indexes are used to take into consideration the effect of inflation on the cost of building treatment plants. The foremost is the Engineering News-Record (ENR) construction cost index (CCI). The other ones are Building Cost Index, which reflects more accurately the types of labor involved in building construction (carpenter, bricklayer, and structural ironworker) and the US Army Corp of Engineers (ACE) Cost Index for Utilities (64). Because of the unique factors involved in treatment plant construction and the lack of an index that accurately portrayed them. US EPA in 1962 developed a specialized index intended to measure the changing cost of treatment plants. Instead of using a theoretical quantity of materials, as the ENR indexes do, the US EPA index chose 20 components representative of a typical high-rate trickling filter plant at 1 MGD ($3{,}785\,m^3/d$) flow. The construction activity was divided into eight major categories, including excavation, process equipment, and buildings. The cost for each category was composed of two or more of the 20 components and their respective prices. When the cost for each category was estimated, the resulting indexes (one prepared for each of the 20 cities used by ENR in their indexes) were averaged to obtain the national index. This index is called the Sewage Treatment Plant Construction Cost Index (STPCCI).

While the STPCCI succeeded in accounting for the special components in treatment plant construction, only wholesale prices were used, as in the CCI. The materials prices that compose the index are lower than those encountered at the retail level. This is particularly evident in an age of high demand and shortages. Furthermore, a trickling filter is no longer a typical type of treatment plant constructed today. This negates the usefulness of the index to some degree. Nevertheless, the STPCCI index is the preferred adjustment factor when available. Accordingly US EPA has prepared guidelines for cost estimates of municipal wastewater systems (65). Unit construction costs of conventional activated sludge plants, extended aeration plants, and contact stabilization plants are presented, among others, in graphical forms. The per capita cost of each treatment process according to design population equivalent taken from the US EPA Guidelines (65) is plotted in Figure 6.16 for comparison. Translated into costs based on flow, the result is presented in Table 6.6. For operation and maintenance costs, Table 6.7 is taken from a report by US EPA (66). To obtain the cost estimates for a particular locality, some regional adjustment is necessary to account for local labor and material costs. The electrical energy consumption and heating requirements of the activated sludge process will affect a plant's operation and maintenance costs significantly. These data can be found in ref. (67).

All cost data presented in Tables 6.6 and 6.7 have been updated to 2008 US Dollar value using US ACE Cost Index for Utilities (64). Unit construction costs shown in Figure 6.16 should be updated to 2008 value by multiplying the 1972 cost by $552.16/141.94 = 3.89$ (see Appendix B).

Table 6.6
Estimated construction cost based on flow

	2008 Construction cost. $ million/MGD			
Type of plant	0.01 MGD	0.1 MGD	1.0 MGD	10.0 MGD
Conventional	12.1	5.83	2.80	1.40
Extended aeration	9.72	5.24	2.60	1.63
Contact stabilization	–	6.22	1.47	–

Table 6.7
Estimated operation and maintenance cost based on flow

	Annual cost (2008 US$) for average daily flow, MGD				
Type of plant	0.1	0.5	1.0	2.5	5.0
Conventional	41,200	108,900	164,500	286,200	435,700
Extended aeration	33,000	78,200	116,400	184,800	–
Contact stabilization	34,300	91,300	148,500	264,900	–

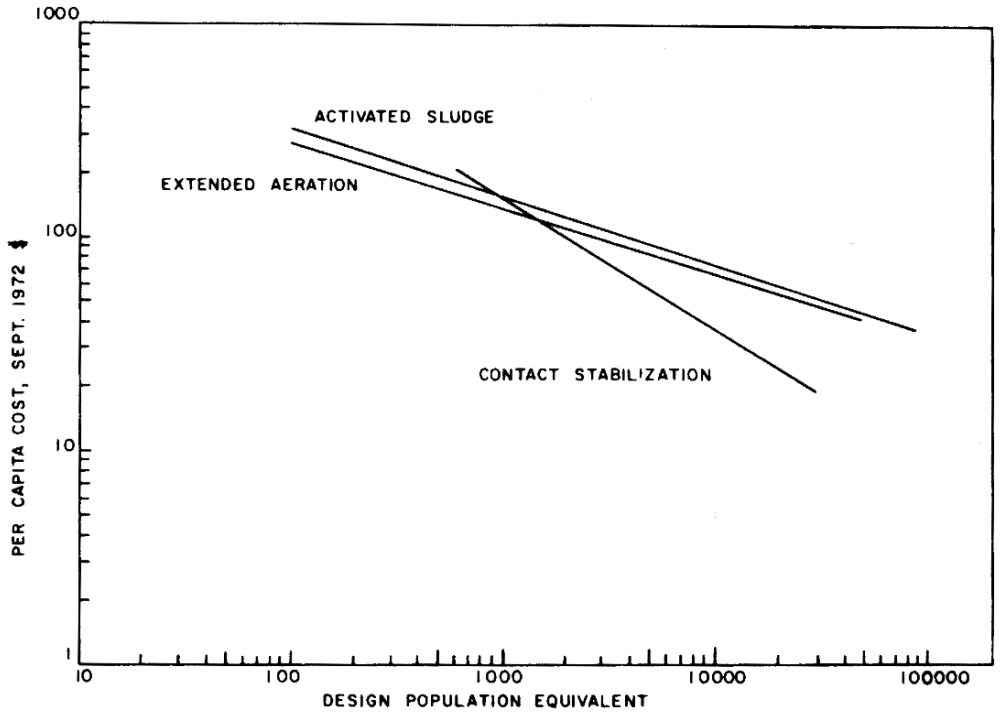


Fig. 6.16. Unit construction costs of activated sludge, extended aeration and contact stabilization plants (65) (To convert cost to 2008 US Dollar value, multiply by 3.89).

6.2. Worksheet for Cost Estimates

The cost estimates given in Section 6.1 are crude at best. They serve the purpose of providing a general comparison of treatment costs for various processes. The major fault in using this general cost information is the fact that biosolids (sludge) treatment and options for reuse/disposal are not considered. This is particularly important for biological treatment processes and some chemical treatment processes from which large quantities of residuals or biosolids are generated (68, 69). The cost of biosolids treatment and disposal in general is 50% or more of the total treatment cost. It is therefore obvious that one should consider several different options, taking into account considerations of land availability, transportation and chemical, power, and fuel rates for cost comparison.

Battelle (70) has made a comprehensive study for the US EPA and the Council on Environmental Quality, of wastewater treatment alternatives with cost estimates. Different treatment strategies are considered, including one for the activated sludge (complete mix) process with discharge to surface water and one for the same process with discharge to land (land application). Each strategy includes eight different biosolids treatment and disposal options at different plant capacities from 1 to 1000 MGD. Profile sheets are provided in the report containing all cost items required for detailed estimates of capital as well as operating costs. A sample worksheet taken from the report (70) to illustrate the proper procedure of cost estimate is included in Appendix A.

6.3. Improvements of Cost Estimation Techniques

In the past, engineers dealt with fixed cost estimates for treatment strategies. Very little attention was given to cost trade-offs in terms of changing demands of effluent quality. In "The Cost of Clean Water" Report (71), for example, industrial waste treatment costs are given as a function of BOD removal efficiencies. A cost function is defined by Tihansky (72) as any relationship that provides an estimate of control costs in terms of the values of more basic variables, called cost determinants or factors. The traditional determinant of unit treatment cost functions for activated sludge processes is the volume or flow rate (73).

In an effort to improve the technique of obtaining a true treatment-cost relationship, Barnard et al. (74) tried to incorporate interrelationships between treatment units into cost estimates. Changes in the operating conditions of certain units are known profoundly to affect the operation of other units in the system. These relationships were used in computer programs developed to determine treatment costs. For treatment of combined municipal and industrial wastewater, using a strategy of pre- and primary treatment followed by activated sludge, sludge thickening, aerobic digestion and sludge dewatering by vacuum filtration with cake disposal, the capital cost model developed by Barnard et al. (74) is as follows:

1. Combined waste treatment

$$\text{Capital cost (\$1,000)} = 500Q^{0.78} + (110 + 37Q)\left(\frac{S_o}{200} - 1\right) + (77 + 23Q)\left(\frac{SS}{200} - 1\right) \quad (24)$$

in which Q = flow rate in MGD, S_o = influent BOD in mg/L and SS = influent suspended solids in mg/L. The effect of removal reaction rate k is not included in the model, but correction factors can be applied.

2. ***Industrial waste***
For treatment of industrial wastewater only, using activated sludge followed by final clarification, thickening aerobic digestion, further thickening, sludge dewatering by centrifugation and cake disposal, but not trucking, Barnard et al. (74) developed the following relationships:

$$\text{Capital cost (\$1,000)} = Q^{w}[17(S_o/S_e)^{0.77} + 215]\left(1.05 + \frac{0.044}{kX_v}\right) \quad (25)$$

$$\text{Capital cost (\$1,000 excluding sludge treatment)} = Q^{u}(S_o/S_e)^{0.4}\left(104 + \frac{10.3}{kX_v}\right) \quad (26)$$

in which,

$$w = 0.69 + 0.0019S_o$$
$$u = 0.66 + 0.000078S_o$$
$$X_v = \text{MLVSS, mg/L}$$

k = reaction rate (l/mg-h) in the equation $S_e/S_o = 1/(1 + kX_v t)$ for activated sludge treatment.

These equations have limited application because they do not allow substitution of other treatment components in the model so that estimates for other treatment alternatives cannot be obtained. Nevertheless, the interrelationships between treatment units in the sets of specified treatment systems and the effects of these relationships on cost estimates have been considered. The technique is unique and represents a new approach to cost formulation. Eilers et al. (75) also developed a computer program for cost estimates using a similar approach and with more flexibility in treatment component substitutions. This approach should be incorporated into future efforts of cost estimates of wastewater treatment plant construction.

Another effort in obtaining better cost estimates of wastewater treatment plants was directed toward developing an index that more accurately reflects the treatment plant construction activities. US EPA, in cooperation with Icarus Corp., developed a new index based on a 50 MGD (189,250 m^3/d) activated sludge treatment plant with some additional tertiary treatment steps. According to a report by JWPCF (76), the model uses over 100 representative components of material and labor combined in representative proportions. These quantities would enter a BICEPS computer program to compute the cost index. The program is used to estimate the cost of the treatment plant in 25 US cities every month and reports the cost of each monthly. Furthermore the index is computed based on actual market prices of materials and labor. Icarus Corp. has a data base that relies on reports from contractors for the price actually paid for items. Also the labor costs were established with the inclusion of productivity figures for the labor skills involved. The US EPA index gives a more realistic picture of the cost of today's treatment plants and has greatly improved the techniques of cost estimates.

7. IMPORTANT DEVELOPMENTS

The ingenuity of wastewater treatment plant operators rather than engineers have in the past developed many modified models of the activated sludge process that are practiced today. Engineers' contributions to the activated sludge process are mainly in research works that lead to the understanding of the process. Kinetic studies and mathematical modeling, using both steady-state and dynamic models, help to identify the major parameters and their influence on the treatment performance. Consequently the process can be better designed and better controlled. Though engineers are busy explaining how the process works, less attention is given to new process development. In this section, many promising process developments will be described (99–133).

7.1. High Rate Adsorption-Biooxidation Process

A unique process using the strong adsorption capacity to enhance biooxidation has been developed by Besik (77). Figure 6.17 is a schematic diagram of the process. Screened wastewater first enters a granular activated-carbon column in expanded bed operation mode. The equilibrium concentration of the colloidal and dissolved organic matter on the activated carbon is much higher than that in the main stream. Consequently for the biomass population adhering to the adsorbent surface, the reaction rates will be also higher. Because activated carbon selectively adsorbs oxygen from an aqueous solution, the oxygen transfer to the cell is also enhanced. The increase in BOD removal rate should be much higher for low organic waste when the cell growth is concentration-dependent.

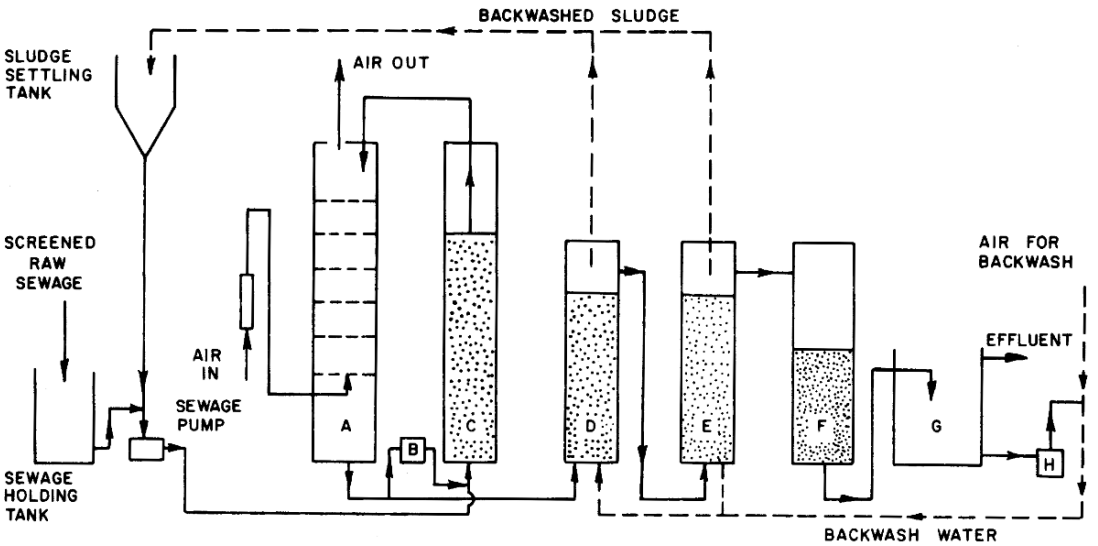


Fig. 6.17. High rate adsorption-biooxidation process (77).

Effluent from the adsorption reactor is aerated in a countercurrent aeration stripping column. A portion of the aerated effluent is recirculated to the adsorption reactor to provide the oxygen needed and the remaining portion is passed through two activated carbon columns and one sand bed for clarification. The activated carbon filters can be backwashed to avoid excessive pressure drop caused by buildup of sludge culture.

For 30-day continuous operation of the pilot plant, using an aeration time from 0.06 to 0.2 hours, volumetric loading of equal to or more than 32 kg BOD/m^3/d ($\geq$ 2000 lb BOD/1000 ft^3/d) based on total adsorption aerator and aeration column volume, the influent with 100 to 300 mg/L TOC can be reduced to less than 3 mg/L of TOC and BOD, turbidity less than three Jackson turbidity units and zero sludge production. The result of zero sludge production cannot be expected in a long term operation. The high potential of the process, however, has been demonstrated. The process seems to be highly efficient and combines the advantages of extended aeration, oxygenated-activated sludge (see Chapter 7) and contact stabilization processes.

7.2. Carrier-Activated Sludge Processes

There has been a substantial interest in recent years in the potential benefits of high biomass wastewater treatment. The major obstacle for achieving this has been the inability of biosolids separation in secondary clarifiers. For the most part, this has been overcome by using various forms of support media or carriers that have the ability to attach high concentrations of aerobic bacterial growth (78–80). The increase in immobilized biomass reduces the process dependence on secondary settling basins for clarification. In such hybrid systems where attached growth coexist with suspended growth one gets more stable systems which possess the combined advantages of both fixed and suspended growth reactors.

7.2.1. Advantages of Biomass Carrier Systems

The performance of carrier systems is dependent on the amount of attached biomass, the characteristics of attached and suspended microorganisms and the type of carriers. The advantages of such hybrid systems are (29):

(a) Heterogeneity of the microbial population. This is brought about by the differences in the microhabitat of organisms attached to the surface of a carrier and those in the bulk of the solution with respect to pH, ionic strength and concentration of organics (81–85).
(b) Increased persistence in reactor. This leads to increase in biomass of organisms, reduction of hydraulic retention time and thus smaller reactor volumes (86–88).
(c) Higher growth rate (89–91).
(d) Increased metabolic activity. This leads to increase in respiration and substrate use, hence higher removal rates (92–95).
(e) Better resistance to toxicity (96–99).

7.2.2. The CAPTOR Process

One interesting concept of hybrid systems is the CAPTOR process developed jointly by the University of Manchester Institute of Science and Technology (UMIST) and Simon-Hartley, Ltd., in the United Kingdom. This high biomass approach uses small reticulated polyurethane

pads as the bacterial growth medium (100). The pads are added to standard activated sludge aeration reactor, and the system is operated without sludge recycle, essentially combining suspended growth with a fixed film in one process. Excess growth is removed from the pads by periodically passing them through specially designed pressure rollers.

The British Water Research Centre (WRC) and Severn-Trent Water Authority conducted a full-scale evaluation of the CAPTOR process for upgrading the activated sludge plant at the Freehold Sewage Treatment Works, in the West Midlands area of England, to achieve year-round nitrification. This full scale study was jointly sponsored by the U.S. Environmental Protection Agency (101, 102).

7.2.3. Development of CAPTOR Process

As mentioned earlier, the CAPTOR process originated from research work on pure systems in the Chemical Engineering Department of UMIST. Single strands of stainless steel wire were woven into a knitted formation and then crushed into a sphere of about 6 mm (0.25 in.) diameter. These particles of known surface area were used for modeling liquid-fluidized bed systems. From this work derived the idea of using porous support pads for growing biomass at high concentrations that could be used in wastewater treatment systems. The idea was jointly developed and patented by UMIST and their industrial partner Simon-Hartley, Ltd. The present form of the CAPTOR process uses 25 mm × 25 mm × 12 mm (1 in. × 1 in. × 0.5 in.) reticulated polyether foam pads containing pores nominally of about 0.5 to 0.9 mm (0.02 to 0.035 in.) diameter and 94% free space (103–105).

7.2.4. Pilot-Plant Study

The conducted pilot-plant work indicated that it was possible to achieve the following (86, 87):

(a) Biomass concentrations of 7000 to 10,000 mg/L.
(b) Waste sludge concentrations of 4% to 6% dry solids using a special pad cleaner.
(c) Improved oxygen transfer efficiencies.
(d) High BOD volumetric removal rates.

7.2.5. Full-Scale Study of CAPTOR and CAST

The full-scale evaluation of the CAPTOR process was undertaken at the Freehold Sewage Treatment Works near Stourbridge, West Midlands. The Freehold plant did not achieve any nitrification in the winter and only partial nitrification in the summer. Freehold's activated sludge system consisted of five trains equipped with tapered fine bubble dome diffusers arranged in a grid configuration. The system was modified as shown in Figure 6.18 to split the wastewater flow into two equal volumes. Half went to two trains that were modified by adding CAPTOR pads to the first quarter of two aeration basins, and the other half went to two trains that remained unaltered and served as a control. The CAPTOR modified trains were each equipped with a CAPTOR pad cleaner (Figure 6.19), and the CAPTOR pads were prevented from escaping into the remainder of the experimental system aeration basins by screens placed at the effluent ends of the CAPTOR zones.

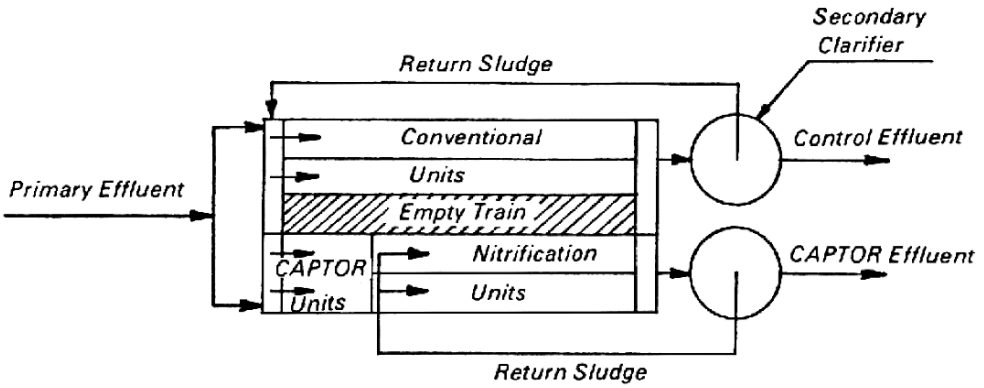


Fig. 6.18. Schematic of treatment plant showing incorporation of CAPTOR (102).

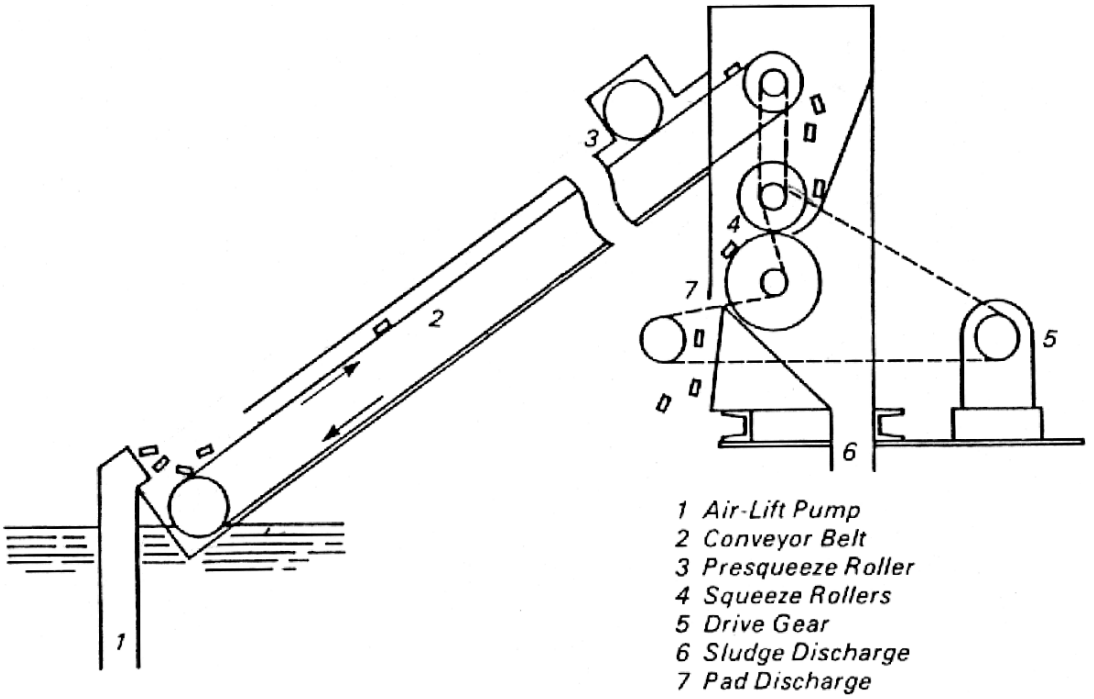


Fig. 6.19. CAPTOR pad cleaner (102).

The Simon-Hartley design predicted that, with a concentration of 40 pads/L, an annual average removal of 75% of the BOD_5 coming into the plant could be achieved in the CAPTOR zones, resulting in a reduced food-to-microorganism (F/M) loading on the follow-on activated sludge stage of 0.08 kg BOD_5/d/kg MLSS. With the reduced load, it was predicted that the modified system would achieve year-round nitrification with an effluent ammonia nitrogen concentration of 5 mg/L or less (102).

7.2.5.1. Full-Scale Plant Initial Results

The Freehold modified CAPTOR activated sludge system was put in operation and immediately encountered a major problem. The CAPTOR pads floated on the surface of the tanks and would not become incorporated into the tank liquor. A solution was found by removing three of the seven longitudinal rows of fine bubble diffusers in the CAPTOR aeration basins. This was done to create a spiral roll in the tanks, which leads to areas of rising and failing liquid with quite large channels down which the pads can fall. The spiral roll modification provided the necessary falling zone and produced complete mixing of the CAPTOR pads.

Another problem that occurred was mal distribution of the pads. The flow of wastewater tended to push the CAPTOR pads to the outlet of their zones, resulting in a concentration of 50 to 60 pads/L at the outlet and only 10 to 20 pads/L at the inlet end.

One other disturbing feature was the rapid deterioration in the CAPTOR pads. The CAPTOR pads used initially were black and were wearing at such a rate that they would not have lasted for more than 3 years, rendering the process uneconomical.

It had also become evident by this time that with the Freehold wastewater it would be possible to achieve the concentration of 200 mg biomass/pad predicted in the design. However, it was found that if the biomass was allowed to grow beyond 180 mg/pad, the biomass in the center of the pad became anaerobic. The control of pad biomass was difficult because the pad cleaners provided were not reliable and were situated at the CAPTOR zone inlets while most of the pads gravitated to the outlet ends of the zones.

During this early period, while the above problems were being tackled on the full-scale plant, there were some occasions when the effluent from the CAPTOR units was reasonable (BOD removals of 40% to 50%), but BOD removal never approached the average of 75% predicted based on the earlier pilot-plant results. Poor BOD removals were being experienced because the suspended solids concentration in the effluent was always high (> 80 mg/L).

Consequently more pilot-scale studies were used to find solutions to the operating problems described above before attempting further full-scale evaluation at Freehold.

7.2.5.2. Pilot-Scale Studies for Project Development

It was decided to evaluate two variations of the CAPTOR process. The new variation differed from the original CAPTOR in that the pads were placed directly into the mixed liquor of the activated sludge aeration tank rather than in a separate stage before the activated sludge tank. WRC named this process variation CAST (CAPTOR in activated sludge treatment). The CAST system had been applied to upgrade several overloaded wastewater treatment plants in Germany and France, and was found to be useful in improving the treatment efficiency and plants performance (106–108).

In addition, a single aeration tank filled with 40 CAPTOR pads/L, was fed effluent from the above activated sludge control unit to assess the potential of CAPTOR as a second-stage nitrification process. Neither pad cleaning nor final clarification was necessary with this process variation because of the low sludge yields characteristic of nitrifier growth.

Studies were conducted using two well-mixed CAPTOR tanks in series. A range of loading and pad cleaning rates were used to evaluate process removal capabilities for CAPTOR. The intermediate effluent was used as a measure of process efficiency of the primary reactor

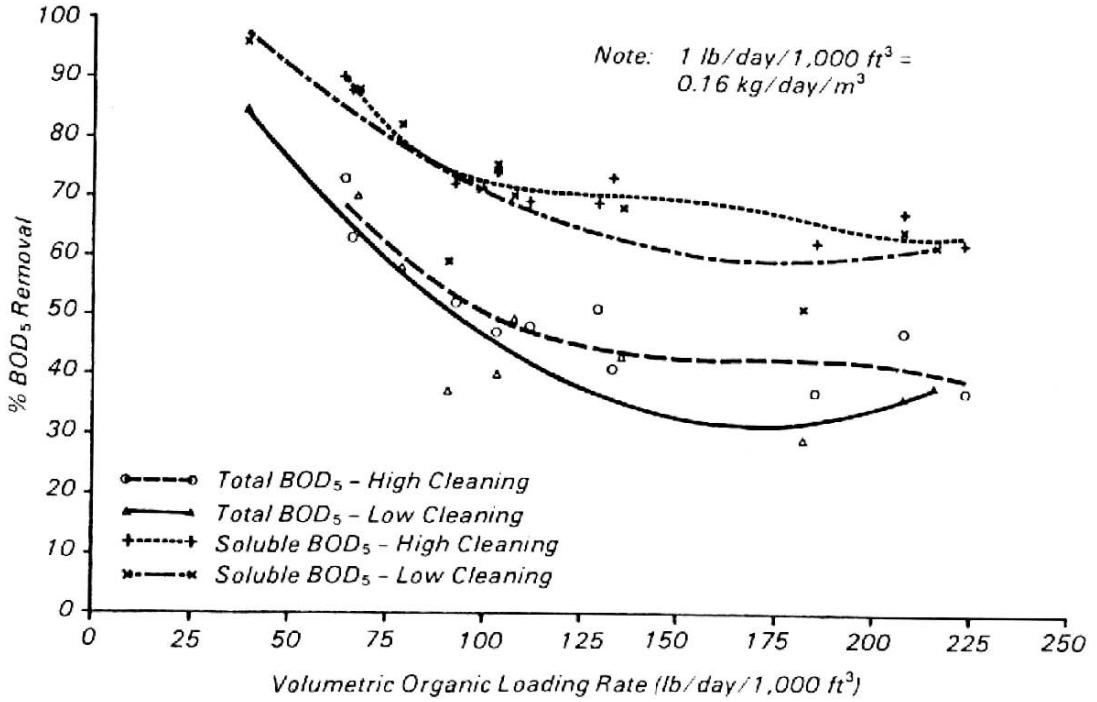


Fig. 6.20. Pilot-scale CAPTOR BOD_5 removals as a function of organic loading rate (102).

and the final effluent for the entire system. This permitted plotting (Figure 6.19) of % BOD_5 removal (total and soluble) vs. volumetric organic loading rate over the range of 1 to 3.5 kg $BOD_5/d/m^3$ (62 to 218 lb/d/1000 ft^3). High and low pad cleaning rates are differentiated in Figure 6.13 as $\geq$16% and <16% of the total pad inventory/d, respectively (102).

Total BOD_5 removal efficiency was less than soluble BOD_5 removal efficiency because of the oxygen demand exerted by the biomass solids lost in the process effluent. The higher pad cleaning rates are believed to have contributed to the improved total and soluble BOD removals shown in Figure 6.20, although low bulk liquid DO's may have adversely affected removals on some of the low cleaning runs. Low cleaning rates (<16%/d) were detrimental to soluble $BO0_5$ removal efficiency because of a gradual decline in activity of the biomass remaining in the pad. Cleaning rates greater than 24%/day, however, resulted in reduced biomass levels in the pads and a reduction in performance.

The problem of mal distribution of CAPTOR pads in the aeration tank (i.e., crowding of pads into the effluent end of the tank when operated in plug flow fashion as at Freehold) was solved by modifying the flow pattern to transverse flow (across the width of the tank rather than down the length). When implemented later at Freehold, this pattern resulted in a fourfold decrease in flow velocity.

Several mixing intensities and diffuser arrangements were tried to decrease biomass shedding into the process effluent. It became obvious; however, that production of effluent biomass

Table 6.8
Pilot-scale operating conditions and process performance (102)

	Period			
Parameter	1		2	
Volumetric loading (lb BOD_5/day/1.000 ft^3)*	113		213	
HRT (hr)	2.32		1.52	
Pads/L	40		40	
Biomass/pad (mg)	121		126	
Equivalent MLSS (mg/L)	4,840		5,040	
FIM loading (kg BOD_5/day/kg/MLSS)	0.37		0.68	
SRT (days)	3.23		1.72	
DO (mg/L)	4.2		4.7	
	In	Out	In	Out
Total BOD_5 (mg/L)	175	93	216	129
Soluble BOD_5 (mg/L)	86	24	85	33
SS (mg/L)	116	120	178	160
Total BOD_5 ramoval (%)	47		40	
Soluble BOD_5 removal (%)	72		61	
SS removal (%)	−3		10	

*1 lb/day/1,000 ft^3 = 0.016 kg/day/m^3

solids was not significantly affected by changes in mixing intensity or diffuser arrangement. High effluent suspended solids proved to be far more dependent on pad cleaning rate, biochemical activity of the biomass, and biomass growth directly in the liquor.

Using the transverse flow scheme and a regular pad cleaning regimen, CAPTOR process performance was similar to that experienced in the small tanks. Operating parameters and process performance are summarized in Table 6.8 for two different volumetric loading rates (102).

Respiration studies conducted on pads indicated that biomass held within the pads respires at up to 40% to 50% less than equivalent biomass in free suspension. Any increase in net biomass concentration achieved in a CAPTOR reactor above that in a conventional activated sludge reactor may not produce noticeable benefits, therefore, owing to the lower specific activity. These observations suggest that diffusion limitations were occurring in the CAPTOR pads.

The CAST variation of CAPTOR was operated in conjunction with a final clarifier to settle the mixed liquor solids component of the total biomass inventory and return it to the aeration tank. CAPTOR pads and biomass retained therein were kept in the reactor by screens. Operating and performance data are compared in Table 6.9 for the CAST unit and the parallel activated sludge control unit for a 25-day period when the volumetric loadings and hydraulic residence times (HRT)s for both units were identical.

In the nitrification experiments conducted on the CAPTOR process, the biomass concentrations per pad ranged from 99 to 124 mg. This is within the range of 100 to 150 mg/L

Table 6.9
Pilot-scale CAST and activated sludge operating conditions and performance (102)

	System			
Parameter	CAST		Activated Sludge	
Volumetric loading (lb BOD_5/day/1,000 ft^3)*	148		148	
HRT (hr)	1.8		1.8	
Pads/L	34		–	
Biomass/pad (mg)	116		–	
Equivalent MLSS in pads (mg/L)	3,930		–	
MLSS in suspension (mg/L)	3,720		6,030	
Total MLSS (mg/L)	7,650		6,030	
F/M loading (kg BOD_5/day/kg total MLSS)	0.31		0.39	
SRT, based on total MLSS (days)	3.6		3.0	
DO (mg/L)	2.5		3.0	
	In	Out	In	Out
Total BOD_5 (mg/L)	178	12	178	20
Soluble BOD_5 (mg/L)	101	5	101	4
SS (mg/L)	121	15	121	23
Total BOD_5 removal (%)	93		89	
Soluble BOD_5 removal (%)	95		96	
SS removal (%)	88		81	

*1 lb/day/1,000 ft^3 = 0,016 kg/day/m^3

reported by other researchers (109). With a pad concentration of 40/L, equivalent MLSS levels varied from 3960 to 4960 mg/L. Liquor DO concentrations were maintained between 6.4 and 8.4 mg/L, and liquor temperature ranged from 11.50 to 6.5°C.

Secondary effluent from the control activated sludge pilot unit used in the CAST experiments was applied to the nitrification reactor over a range of loading conditions. Essentially complete nitrification was achieved at TKN and ammonia nitrogen loadings of approximately 0.25 kg/d/m^3 (15.6 lb/d/1,000 ft^3) and 0.20 kg/d/m^3 (12.5 lb/d/1000 ft^3), respectively.

7.2.5.3. Full-Scale Plant Results after Modifications

Following the successful testing of the transverse mixing arrangement in the pilot-scale study, the two Freehold CAPTOR trains were modified. The modifications involved the following (102):

(a) Splitting each of the CAPTOR trains, C1 and C2, into two compartments, C1A and C1B and C2A and C2B, as shown in Figure 6.21.
(b) Feeding influent flow along long weirs at the side of the trains instead of at the narrow inlet ends.
(c) Modifying the aeration pipe work to place all three rows of dome diffusers directly below the outlet screens (covering about 25% of the width of the tanks), thereby creating a spiral roll of pads and liquid counter-current to the flow of wastewater entering along the weirs on the sidewalls.
(d) Installing two extra pad cleaners so that each CAPTOR sub-unit was provided with a cleaner.

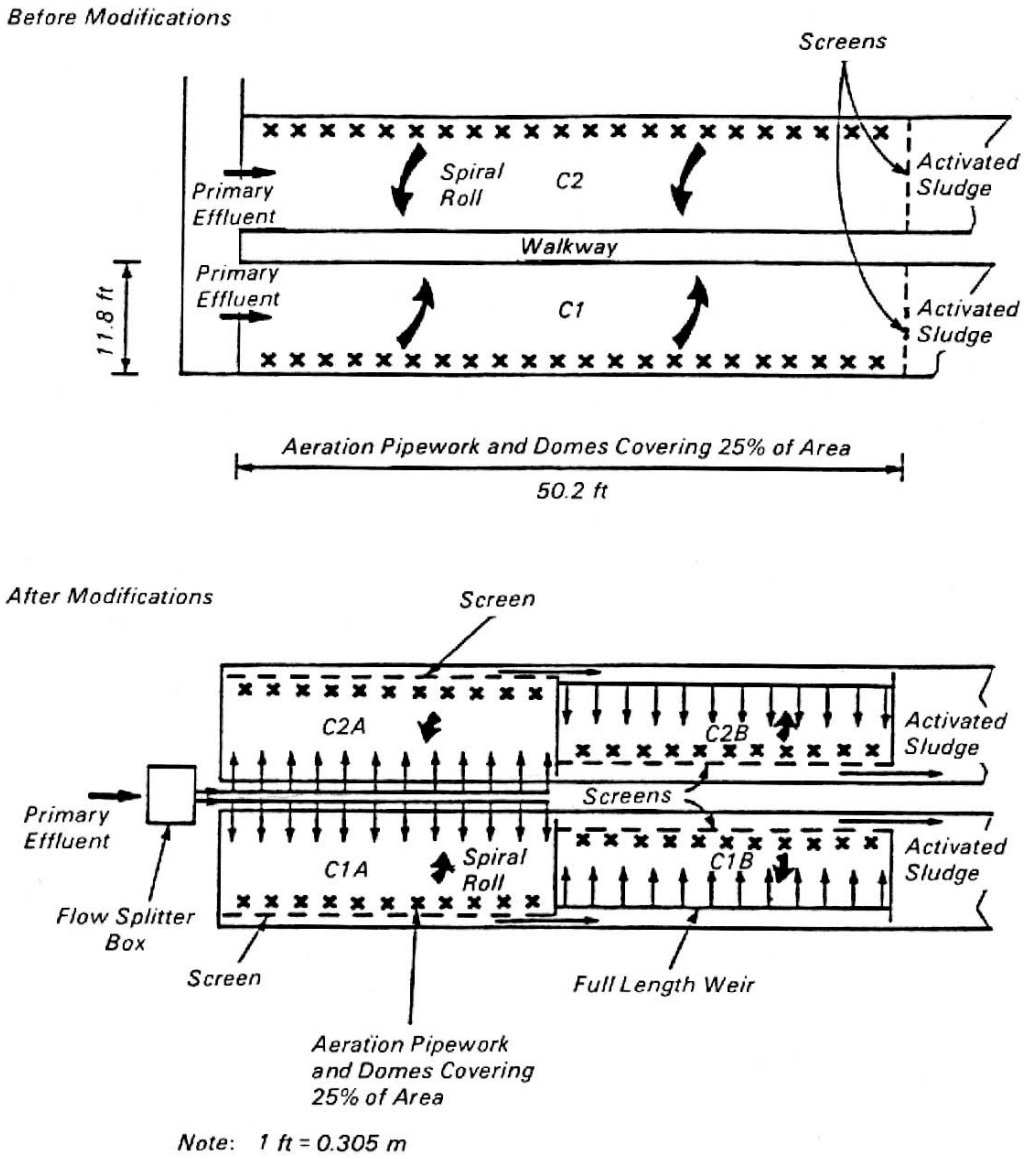


Fig. 6.21. Modifications to full-scale CAPTOR system flow pattern (102).

(e) Installing fine screens at the outlet from the primary clarifiers to reduce the quantity of floating plastic material entering the CAPTOR units that created problems with the cleaners.

The objective of the first three modifications was to achieve uniform mixing of the pads in the CAPTOR units and prevent the situation that had occurred previously where high concentrations of pads (50 to 60 pads/L) collected at the outlet end and very low concentrations (10 to 20 pads/L) at the inlet end. Pads were removed from the tanks during the modifications. After

Table 6.10
Full-scale modified CAPTOR performance results (102)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD_5	128	22	83
Soluble BOD_5	40	4	90
SS	138	32	77
NH_4-N	24	24.4	0

Table 6.11
Full-scale modified CAST performance results (102)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD_5	138	16	88
Soluble BOD_5	56	2	96
SS	120	27	78
NH_4-N	26.7	17.2	36

the modifications were completed, the number of pads in each compartment was equalized at about 35/L.

The changes were completely successful in obtaining uniform distribution and complete mixing of the CAPTOR pads. A lithium chloride tracer test conducted on the modified tanks indicated that no dead zone was occurring in the "eye" of the roll. Formation of floating pad rafts (which had occurred at the outlet end of the tank with the original arrangement) was completely eliminated. The modifications, however, had no effect on the high level of suspended solids present in the liquor. The modified CAPTOR system was operated at an average volumetric loading rate of 1.24 kg $BOD_5/d/m^3$ (77 $lb/d/1000\,ft^3$), an average HRT (excluding sludge recycle) of 2.55 hours and an overall biomass concentration of 4,830 mg/L.

The CAST variation of the CAPTOR process, which had exhibited somewhat better performance than conventional activated sludge in the small tank experiments, was also field evaluated at Freehold. The CAPTOR trains were further modified so that return sludge could be introduced to the CAPTOR zones (35 pads/L), providing an activated sludge component throughout the entire aeration tanks, not just in the nitrification stage. The average volumetric organic loadings and HRTs (excluding sludge recycle) were 1.11 kg $BOD_5/d/m^3$ (69 $lb/d/1000\,ft^3$) and 3.40 hours, respectively.

Performance data summarized in Tables 6.10 and 6.11 indicate that the CAST system exhibits somewhat better performance than the CAPTOR version. In the CAST process the removal of soluble BOD_5 is 96% compared to 90% in CAPTOR; the removal of total BOD_5 is 88% compared to 83%; and the removal of SS is about the same at about 78%

7.2.5.4. Overall Conclusions

The US EPA conclusions and recommendations for the CAPTOR/CAST treatment systems are as follows (101, 102, 110):

In the initial phase when the CAPTOR process was installed at the Freehold Sewage Treatment Works, several problems were immediately evident. There were major problems with respect to pad mixing, suspension, and distribution and the process performance was adversely affected by the high level of suspended solids in the CAPTOR stage effluent. The problems of pad mixing and distribution were solved by pilot- and full-scale development work

(a) The performance of the CAPTOR process was still adversely affected by the high level of suspended solids in the CAPTOR stage effluent after correction of the pad mixing, suspension, and distribution problems. This prevented the achievement of nitrification in the follow-on activated sludge stage.
(b) The presence of CAPTOR pads in the tank liquid did not improve oxygen transfer efficiency.
(c) The durability of the CAPTOR pads was solved by switching to different pads.
(d) The peak biomass concentration in the pads is unpredictable. It does not appear to be related to the BOD concentration of the wastewater. There were indications in the various studies, however, that the frequency of pad cleaning (and, hence, the biomass/pad concentration) was critical to the performance of the process. Regular pad cleaning is essential to prevent anaerobic conditions from developing in the pads.
(e) It is possible to raise the biomass concentration in a CAPTOR stage to 6000 to 8000 mg/L, but the respiration rate of the biomass in the pads is lower than the respiration of the same biomass if freely suspended and less than that of normal activated sludge. These data suggest that the geometry of the CAPTOR pads results in diffusion limitations, which demands further pad design improvement to enhance the potential for economic use of the CAPTOR process in wastewater treatment.
(f) The CAST variation of the CAPTOR process performs well.
(g) CAPTOR has the potential as an add-on package for tertiary nitrification.
(h) The CAPTOR option was projected to be more cost effective than extending the activated sludge plant for upgrading Freehold to complete year-round nitrification.
(i) For CAPTOR and CAST to achieve their full potential, as predicted by the pilot-scale studies, further design development and improvements are needed.

7.3. Secondary Flotation Process

The most common operational difficulties encountered in the conventional biological treatment plant are rising sludge and bulking sludge, resulting in high suspended solids and BOD in the plant effluent. The common causes and efficient remedies were discussed earlier in this chapter. In summation there are two principal types of sludge bulking problems: (a) the growth of filamentous organisms, (b) the formation of swelling biological flocs through the addition of bound water to the cells to the extent that their density is reduced. Possible causes of sludge bulking include:

(a) Absence of certain necessary trace elements in wastewater.
(b) Wide fluctuations in wastewater pH.
(c) Limited dissolved oxygen in the aeration tank.
(d) Inadequate food-to-microorganism ratio.
(e) Inadequate mean cell residence time.
(f) Inadequate return sludge pumping rate.
(g) Internal plant overloading.
(h) Poor sedimentation clarifier operation.

The problems of rising sludge and sludge bulking, when serious, cannot be overcome easily. If rising and bulking conditions continue to persist after all the aforementioned factors have been checked, a critical investigation of the behavior of aeration basin and secondary sedimentation clarifier should be made. It is very possible that the design is at fault, and either changes or expansions must be made in facilities.

Expansions in the existing aeration basins and secondary sedimentation clarifiers are costly and sometimes unaffordable. The easiest facility upgrade will be the addition of a secondary flotation unit, which is primarily a dissolved air flotation clarifier (1, 57, 111, 112).

The secondary sludge in the secondary flotation clarifier is floating, thus sludge rising is no longer a problem, in fact, becomes a big plus.

For an existing activated sludge treatment plant, a secondary flotation clarifier can be applied in series between the aeration basin and secondary sedimentation to separate the living microorganisms before settling. This results in the following improvements: (a) Solids and hydraulic loadings on the overloaded secondary sedimentation are reduced for preventing sludge rising, increasing clarification efficiency, and saving construction cost on expansion of secondary sedimentation facilities; (b) Hydraulic loading on the aeration basin is reduced thus increasing retention time without increasing aeration basin size; (c) Higher solids content of the waste sludge represents cost saving and improved operation of sludge thickening, dewatering and disposal; (d) The return sludge, separated by flotation stay in aerobic condition at all times and is more active than comparable settled activated sludge. The oxygen requirement for the mixed liquor suspend solids is also reduced owing to the recycling of aerobic sludges, and (e) The problems of sludge rising and sludge bulking can be totally solved when using flotation (57).

For design of a new activated sludge treatment plant, either secondary flotation (57) or a newly developed sandfloat process (111, 112) could be considered as a replacement of secondary sedimentation for the elimination of possible sludge bulking and sludge rising problems. Sandfloat is an improved secondary flotation clarifier with a built-in tertiary filter. If the sandfloat process is to replace the secondary sedimentation in an activated sludge process system, it will return the floated activated sludge to the biological reactor in an aerobic condition, enhance nitrification, remove phosphate and fungi, increase the detention time of biological reactor, and save the land space for construction. The quality of sandfloat effluent will also far exceed the required secondary effluent standards.

The improved activated sludge process system using secondary flotation may accomplish two objectives: (a) to obtain the maximum possible removal of organic substances with the shortest possible time using the best biological flocs that normally settle poorly but float efficiently, and (b) to produce and recycle flocculant biological flocs having a good floating characteristic. The two objectives are compatible.

7.4. Nitrification and Denitrification

Owing to the stricter effluent standards, in particular for nutrients, many existing wastewater treatment plants need to be upgraded. Upgrading of plants with conventional processes may require the extension of tank volumes and the construction of new reactors. It is necessary to develop innovative processes, which satisfy the following requirements: high removal

efficiencies, upgrading with minimal interference with the existing facilities, low investments, and simple technologies with low operating costs (113).

Conventional activated sludge treatment for nitrogen removal include several phases with different oxygen concentrations or may employ only one reactor in which alternating aerobic and anoxic phases are achieved in time or space. In such processes, the oxygen concentration must be controlled to obtain the simultaneous (in time and in space, i.e., with constant conditions and in the same reactor) activity of denitrifying and nitrifying bacteria. It has been shown, in fact, that denitrification reaction takes place also under aerobic conditions and that nitrification is possible at low levels of dissolved oxygen. As far as denitrification is concerned, "aerobic denitrification" and "microzones denitrification" can be defined (114). In the former, micro-organisms use nitrate and oxygen simultaneously as terminal electron acceptors; and in the latter, denitrification occurs in the anoxic microzones with a biological floc. This last phenomenon is supposed to occur owing to the particular aeration conditions at very low dissolved oxygen concentration.

Low oxygen concentrations can be maintained by direct DO control or indirect DO control, like oxidation-reduction potential (ORP). Because nitrification and denitrification have become major aims in wastewater treatment, attempts have been made to use ORP as a control parameter, cheap but reliable control strategies, for wastewater treatment plants (115). Most ORP controllers based on curve interpretation, and those also have implemented time-control strategies as an emergency control whenever characteristic points cannot be found by controller (116).

Other studies for the nitrification-denitrification process include nitrification performance in an integrated fixed activated sludge process (117), and iterative design of a nitrate controller using an external carbon source in an activated sludge process (118). In an activated sludge process, the nitrogen removal efficiency may be improved by adding an external carbon source. Automatic control of the nitrate level can be achieved by regulating external carbon flow (119–121).

7.5. Membrane Bioreactor

In recent years, new membranes specially developed for use in wastewater treatment have made membrane bioreactors application become a promising alternative to the well-known aerobic processes. Ultra- or micro-filtration membrane possesses advantage in possible replacing conventional sedimentation for the separation of the treated water from the sludge (122). The use of submerged membranes has reduced the power consumption of membrane bioreactors and hence increased their potential for the application of membranes in wastewater treatment. Moreover, ultra- or micro-filtration membranes with a pore-size of 0.2 μm or less not only retain bacteria but also viruses (123). The complete retention of sludge allows operation at much higher biomass concentrations. The higher the concentration the lower the *F/M* ratio becomes, with the effect that the microorganisms use a growing portion of the carbon content of the feed for maintenance.

Urbain, et al. (124) and Rosenberger, et al. (125) demonstrated the good performance of aerobic treatment of municipal wastewater using a membrane bioreactor. Rosenberger, et al. (125) pilot plant comprised an anoxic zone to enable denitrification. The hydraulic retention

time (HRT) varied between 10.4 and 15.6 hours.; accordingly, the volumetric loading rate was between 1.1 and 1.7 kg COD/m^3/d. The mixed liquor suspended solids concentration gradually increased to 18 to 20 g MLSS/L. The F/M ratio varied according to the operation conditions but decreased to a value of 0.07 kg COD/kg MLSS/d. Treatment performance was very stable and on a high level. The COD was reduced by 95%. Nitrification was complete and up to 82% of the total nitrogen could be denitrified.

7.6. Reduction of Excess Sludge

It has been estimated that handling, treatment and disposal of a large amount of excess waste sludge, produced daily in a typical municipal wastewater treatment plant, accounts for 50% to 60% of the operating cost of a secondary treatment plant (126). In addition, its ultimate disposal by landfill and/or incineration has created environmental challenge because the availability of landfill sites and incineration of solid wastes poses great difficulties in densely populated nations. Against such a background, attempt was made to use ozone gas to dissolve the excess sludge (127), thereby leading to 100% minimization of excess sludge within the process. In this approach, a small amount of return sludge is ozonated and then returned to the aeration tank. It has been demonstrated that sludge ozonation can render the excess sludge biologically oxidized. Recent studies (68, 128–130) also proved that sludge ozonation treatment is a potential solution to the excess sludge problem at a full-scale operation.

Saby, et al. (131) applied a chlorination step to reduce excess sludge in activated sludge process. It is a strong oxidizer and the chlorination operation cost is only 10% of that of ozonation in terms of disinfection. The chlorination of excess sludge at a chlorine dose of 0.066 g Cl_2/g MLSS and a retention time of 20 hr followed by the recycling of sludge to the activated sludge system resulted in a 65% reduction of excess sludge. The principal disadvantage of the sludge chlorination is the formation of trihalomethanes (THMs). It was found that less than 200 ppb THMs were detected in the effluent of the system. Thus, the THMs formation in water did not become an issue in the process; nevertheless, it seemed necessary to investigate the gas emission during the chlorination step (132).

8. DESIGN EXAMPLES

8.1. Design Example 1

Wastewater characteristics

Primary effluent having a BOD of 200 mg/L and negligible VSS,
Growth yield coefficient = 0.5,
$k_m = 1.39$/d;
$K_s = 50$ mg/L
BOD, endogenous or decay coefficient = 0.1/d,
Q, rate of flow = 3,785 m^3/d (1.0 MGD).

Specified design criteria:

Mixed liquor volatile suspended solids (MLVSS) = 3000 mg/L
Sludge retention time or sludge age = 8 days

Effluent BOD = 30 mg/L
Effluent VSS = 10 mg/L (within US EPA limits of secondary effluent quality).

Assumed conditions:

MLVSS = 80% MLSS
Return sludge = 10,000 mg/L SS or 8000 mg/L VSS
60% of effluent VSS are biodegradable

Solution: design for a complete-mix model

Effluent substrate concentration, S,

$$\begin{aligned} S &= \text{effluent BOD} - \text{BOD of effluent biological solids} \\ &= 30 - 10\,(0.6)(1.42)(0.68) \\ &= 24\,\text{mg/L soluble BOD} \end{aligned}$$

The calculation assumes 1.42 g O_2 required for every gram of VSS (i.e. biological solids) decayed according to the reaction

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3$$

and 68% of ultimate BOD is satisfied in 5 d when the decay rate $b = 0.1/\text{d}$

Biological treatment efficiency, E

$$E = \frac{(200 - 24)}{200}(100) = 88\%$$

Overall plant efficiency, E_{overall}

$$E_{\text{overall}} = \frac{(200 - 30)}{200}(100) = 85\%$$

Aeration tank volume V,

Rewrite Equation (10) as

$$XV = \frac{\theta_c QY(S_o - S)}{1 + b\,\theta_c}$$

or

$$\begin{aligned} V &= \frac{\theta_c QY(S_o - S)}{X(1 + b\,\theta_c)} \\ &= \frac{8(3{,}785)(0.5)(200 - 24)}{3{,}000[1 + 0.1(8)]} = 495\,\text{m}^3 \quad (1.3 \times 10^5\,\text{gal}) \end{aligned}$$

Hydraulic retention time θ,

$$\theta = \frac{V}{Q} = \frac{495}{3{,}785} = 0.13\,\text{d} \quad (3.1\,\text{h})$$

Recalculation of R,

Balancing the mass of solids entering and existing in the aeration tank,

$$3000\,(Q + Q_r) = 8000\,(Q_r)$$

$$R = Q_r/Q = 0.6$$

Specific substrate use rate, U,

$$U = \frac{dS/dt}{X} = \frac{(200-24)(3{,}785)(10^3)}{3{,}000(495)(10^3)} = 0.45/\text{d}$$

Volumetric loading (process loading)

$$\text{Loading} = \frac{(3785)(10^3)(200-24)}{495(10^6)} = 1.35\,\text{kg BOD/m}^3\text{/d} \quad (84.4\,\text{lb BOD/1000 ft}^3\text{/d})$$

Wasted sludge flow, Q_w

Assuming 2.5 hour retention time for the final clarifier and therefore volume V_s of the clarifier $= 395\,\text{m}^3$

$$Q_w = \frac{(V + V_S)X}{X_r\,\theta_c} = \frac{(495+395)(3{,}000)(10^3)}{8{,}000(10^3)(8)} = 42\,\text{m}^3\text{/d} \quad (1.11 \times 10^{-2}\,\text{MGD})$$

Sludge production rate,

$$Q_w X_r = \frac{42(10^3)\,(8{,}000)}{10^6} = 336\,\text{kg VSS/d or } 420\,\text{kg SS/d}$$

Solution: design for a plug flow model

To obtain the same treatment efficiency, a shorter solid retention time is required for the plug flow model, using Equation (11a):

$$\frac{1}{\theta_c} = \frac{Y k_m (S_0 - S_1)}{(S_0 - S_1) + K_s(1+R)\ln\left[\frac{RS_1+S_0}{(1+R)S_1}\right]} - b$$

$$= \frac{0.5(1.39)(200-24)}{(200-24) + 50(1+0.6)\ln\left[\frac{0.6(24)+200}{(1+0.6)(24)}\right]} - 0.1$$

$$\theta_c = 3.45\,\text{d}$$

Aeration tank volume, V

$$V = \frac{\theta_c Q Y (S_0 - S_1)}{X(1 + b\theta_c)}$$

$$= \frac{3.45(3785)(0.5)(200-24)}{3000[1 + 0.1(3.45)]} = 285\,\text{m}^3$$

Hydraulic retention time, θ

$$\theta = \frac{V}{Q} = \frac{285}{3785} = 0.075\,\text{d or } 1.8\,\text{hr}$$

Specific substrate use rate U,

$$U = \left[\frac{dS}{dt}\right] \Big/ dX = \frac{(200-24)(3785)(10^3)}{3000(285)(10^3)}$$
$$= 0.78\,d^{-1}$$

Volumetric loading (process loading)

$$\text{Loading} = \frac{(3785)(10^3)(200-24)}{285(10^6)}$$
$$= 2.35\,\text{kg BOD/m}^3\text{/d or}146\,\text{lb BOD/1000 ft}^3\text{/d}$$

Wasted sludge flow, Q_w

$$Q_w = \frac{(V+V_s)\bar{X}}{X_r\,\theta_c} = \frac{(285+395)(3{,}000)(10^3)}{8{,}000(10^3)(3.45)} = 74\,m^3/d \quad (0.02\,\text{MGD})$$

Sludge production rate

$$Q_w X_r = \frac{74(10^3)(8{,}000)}{10^6} = 590\,\text{kg VSS/d} \quad (740\,\text{kg SS/d})$$

For the sake of comparison, if the same aeration tank volume used in the complete-mix model is used for the plug flow model with sludge retention time = 8 days, the effluent substrate concentration S_1 can be calculated from Equation (11a),

$$\frac{1}{\theta_c} = \frac{Yk_m(S_0-S_1)}{(S_0-S_1)+K_s(1+R)\ln\left[\frac{(RS_1+S_0)}{(1+R)S_1}\right]} - b$$

$$\frac{1}{8} = \frac{0.5(1.39)(200-S_1)}{(200-S_1)+50(1+0.6)\ln\left[\frac{(0.6S_1+200)}{(1+0.6)S_1}\right]} - 0.1$$

By trial and error, solving for S_1

$$\text{Effluent soluble BOD, or } S_1 = 1.0\,\text{mg/L}$$

A biological treatment efficiency of $[(200-1)/(200)](100) = 99.5\%$ is obtained compared to 88% for a complete-mix mode operation.

The example demonstrates that a plug flow system theoretically is more efficient in removing the soluble BOD than a complete-mix system. The difference of treatment efficiency is insignificant for obtaining a treated effluent of 20 mg/L BOD or above. Nevertheless the higher efficiency of treatment in the plug flow mode operation is obvious for secondary treatment of most wastewaters.

8.2. Design Example 2

Determine the air requirement in the operation of a complete mix activated sludge process described in Example 1.

Solution

1. Theoretical oxygen demand, O_2

$$\text{BOD}_\text{L} \text{ of wastewater to be removed} = \frac{(200-24)(10^3)(3{,}785)}{0.68(10^6)} = 980\,\text{kg O}_2\text{/d}$$

$$\text{O}_2 = 980 - 1.42 \text{ (daily VSS production)}$$

$$= 980 - 1.42(336)$$

$$= 503\,\text{kg O}_2\text{/d}$$

2. Volume of air required, V_a
Assume air contains 21% oxygen by weight and its specific weight is 1.2 kg/m^3(0.075 lb/ft^3)

$$\text{O}_2 = \frac{503}{1.2(0.21)} = 1{,}996\,\text{m}^3\text{/d} \quad (7.05 \times 10^4\,\text{ft}^3\text{/d})$$

If the oxygen transfer efficiency is 10%,

$$V_\text{a} = \frac{1{,}996}{0.1} = 19{,}960\,\text{m}^3\text{/d} \quad (490\,\text{cfm})$$

3. Volume of air per unit mass of BOD removed

$$\frac{19{,}960}{(200-24)(10^3)(3{,}785)(10^{-6})} = 30\,\text{m}^3\text{/kg BOD removed} \quad (482\,\text{ft}^3\text{/lb})$$

4. Volume of air per unit volume of wastewater treated

$$\frac{19{,}960}{3{,}785} = 5.3\,\text{m}^3\text{/m}^3 \text{ wastewater} \quad (0.71\,\text{ft}^3\text{/gal})$$

5. Horsepower required if mechanical aeration is used
Assuming each aerator has a capacity to deliver 1.5 kg O_2/hp/hr

$$\text{hp required} = \frac{503}{1.5 \times 24} = 14\,\text{hp}$$

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NOMENCLATURE

α = Specific consumption of oxygen, mass/volume/unit time
b = Endogenous respiration or decay coefficient, time-1
C_o = DO concentration of the floc surface, mass/volume
C_i = DO concentration in the innermost cell of floc, mass/volume
D = Dilution rate, reciprocal of hydraulic retention time
D = Diameter of a sphere, cylinder or the thickness of a plane or parallel floc

E = Electrode potential, millivolts
E_o = Electrode potential when [oxidant] = [reductant], millivolts
E_a = Activation energy, cal/mol
F = Faraday, 23,060 cal/eV
F/M = Food to microbial mass ratio, mass of substrate applied for unit biomass per day
ΔG = Gibbs free energy, kcal/electron-mol
K_a = Adsorption removal constant
k_m = Maximum rate of specific substrate use, time-1
k_1 = Reaction rate at temperature T_1
k_2 = Reaction rate at temperature T_2
K_{20} = Cell growth rate or substrate use rate at 20°C
K_s = Michaelis-Menten constant, or half-velocity coefficient, numerically equal to the substrate concentration when $U = 1/2\,k$. mass/volume
K_T = Cell growth rate or substrate use rate at temperature T
n = Number of electron moles transferred per mole of substrate used for energy, or station number downstream of each subaerator
θ = Hydraulic retention time
θ_c = Sludge retention time, sludge age, or mean cell residence time
θ_T = Temperature coefficient, a constant
Q = Flow rate, volume/time
Q_r = Recirculated flow rate, volume/time
Q_w = Wasted sludge flow rate, volume/time
R = Recirculated flow ratio, Qr/Q
S = Substrate concentration, mass/volume
S_{ao} = Initial concentration of organics removable by adsorption, mass/volume
S_a = Concentration of remaining organics removable by adsorption, mass/volume
S_1 = Effluent substrate concentration of the plug flow process, mass/volume
S_o = Initial substrate concentration, mass/volume
t = Time
U = Specific substrate use rate, change of soluble substrate concentration per unit time per unit microbial mass, time-1
μ = Net specific growth rate, change of microbial concentration per unit time per unit microbial concentration, time^{-1}
V = Volume
V_S = Volume of final clarifier
X = Microbial concentration, mass/volume
$\bar{X}$ = Average microbial concentration in the plug flow process, mass/volume
X_c = Sludge concentration in the final clarifier effluent, mass/volume
X_θ = Sludge production rate for a complete mix model with no sludge recycle, mass/time
$X_{\theta c}$ = Sludge production rate for a complete mix model with sludge recycle, mass/time
X_r = Return sludge concentration, mass/volume
X_v = MLVSS, mixed liquor volatile suspended solids, mg/L
Y = Growth yield coefficient, mass microbial growth per unit mass substrate used

DEFINITION OF TERMS; CASSO PROGRAM

BE = Final effluent 5-d BOD concentration, mg/L
Bi = Primary effluent 5-d BOD concentration, mg/L
BL = BOD loading, *WTBI/WTML,* lb BOD applied/lb MLSS in the aeration tank
F = Wastewater flow, MGD
Q = Unit rate of BOD removal, (*WTBJ-WTBE*)/(*WTML*). lb BOD removed d/lb of MLSS in aeration tank
R = Return sludge flow, MGD
RT = Hydraulic retention time in aeration tank, $V/(F + R)$, d
S = Surface area of sedimentation tank, ft^2
SA = Sludge age, $(WTML)/(WTSW + WTSE)$, d
SE = Final effluent SS concentration, mg/L
SI = Primary effluent SS concentration, mg/L
SL = SS Loading, *WTSI*/*WTML*, lb SS applied/d/lb MLSS in aeration tank
SOR = Surface overflow rate of sedimentation tank, $(F + R)/S$, gpd/ft^2
T = Wastewater temperature, °F
V = Volume of aeration tank, million gallons or MG
W = Waste sludge flow, MGD
$WTBE$ = Weight of BOD in final effluent, *(F − W)(BE)(8.34)/2000,* tons/d
$WTBI$ = Weight of BOD in primary effluent, $(F)(BE)(8.34)/2000$, tons/d
$WTML$ = Weight of MLSS in aeration tank, $(V)(X)(8.34)/2000$, tons
$WTSE$ = Weight of SS in primary effluent, $(F - W)(SE)(8.34)/2000$, tons/d
$WTSI$ = Weight of SS in primary effluent, $(F)(SI)(8.34)/2000$, tons/d
$WTSO$ = Weight of MLSS in the aeration tank effluent, $(F + R)(X)(8.34)/2000$, tons/d
$WTSR$ = Weight of return sludge SS, $(R)(YR)(8.34)/2000$, tons/d
$WTSW$ = Weight of waste sludge, $(W)(YR)(8.34)/2000$, tons/d
X = MLSS concentration, mg/L
Y = Sludge yield coefficient, $(WTSW + WTSE)\,I\,(WTBI - WTBE)$, lb sludge formed/lb BOD removed
YR = Return sludge as concentration, mg/L

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APPENDIX A

Sample Worksheet for Cost Estimates (70)

Assumptions

Plant type:	
Flow rate:	
Prevailing construction index:	(STPCCI) US EPA: ENR: ACE
Local multiplier:	
Local land value:	$ /acre
Local cost of labor:	$ /hr (weight average for operating labor and supervision)
Local cost of power:	$ /kwh
Prevailing cost of chemicals:	
	_____@ $_____/lb
	_____@ $_____/lb
	_____@ $_____/lb
	_____@ $_____/lb
Transport distance for sludge:	miles round trip
Transport cost for sludge:	$/ton-mile
Transport distance for effluent:	miles
Transport cost for effluent:	$/1,000 gal
Amortization basis:	_____%
	_____Years
Factor	_____

Capital Costs	***Estimated Cost***
1. Running total base cost on profile sheet $ _____ for _____ flow	
2. Flow change multiplier = (New flow rate/Old flow rate probile sheet)$^{0.6}$	= _____
3. National cost factor = ( )/175 (STPCCI)	= _____
Price index adjusted cost	= $_____
4. Local multiplier	= _____
5. Total adjusted cost	= $_____
6. Base land requirement from profile sheet	_____ acres
7. Land requirement multiplier = (__/__)$^{0.6}$	= _____
8. Adjusted land requirement = _____ × _____	= ___acres
9. Adjusted land costs = _____ × _____	= $_____
10. Total capital expenditures:	= $_____

Operating Costs

1. Base labor requirement on profile _____ person years
2. Labor multiplier (___/___)$^{.58}$ = _____person years
3. Adjusted labor requirement = _____ person years
4. Base electrical requirement on profile sheet _____ kwh/d
5. Electrical multiplier (___/___).55 = _____
6. Adjusted electrical equipment = _____ × _____ = _____kwh/d
7. Base chemical requirements on profile sheet _____ lb/d
8. Chemical multiplier = (___/__)$^{1.0}$ = _____
9. Adjusted chemical requirements = _____ × _____ = _____ lb/d
10. Base fuel requirement on profile sheet _____ Btu/d
11. Fuel multiplier = (___/____)$^{1.0}$ = _____
12. Fuel requirement = _____ × _____ = _____Btu/d
13. Base quantity of sludge requiring transportation _____ tons/d
14. Sludge multiplier (___/__) = _____
15. Adjusted quantity of sludge requiring transportation _____ tons/d
16. Volume of effluent requiring transportation _____ 1,000 gal/d
17. Daily labor cost _____ × _____ × 8 = $_____/d
18. Daily electrical cost _____ × _____ = $_____/d
19. Daily chemical cost _____ × _____ = $_____/d
20. Daily fuel cost _____ × _____ = $_____/d
21. Daily solids transportation cost_____ × _____ = $_____/d
22. Total daily cost _____ + _____ + _____ + _____ + _____ = $_____/d
23. Total cost/1000 gal _____ = _____/1000 gal
24. Cost for effluent transportation _____ × _____ × _____ = $_____/1000 gal
25. Adjusted operating costs _____ + _____ = $_____/1000 gal
26. Adjusted amortization cost at 10% for 20 yr (__ × _____)(365 × ___)(365 × ___) = $_____/1,000 gal
27. Total adjusted operating cost _____ + _____ = $_____/1000 gal

APPENDIX B

United States Yearly Average Cost Index for Utilities U.S. Army Corps of Engineers (64)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

7
Pure Oxygen Activated Sludge Process

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract The use of pure oxygen for activated sludge treatment has become competitive with the use of air owing to the development of efficient oxygen dissolution systems. The pure oxygen system may be used for aeration in activated sludge systems that operate in either the plug flow or complete mix hydraulic regimes. It is readily adaptable to new or existing complete mix systems and can be used to upgrade and extend the life of overloaded plug-flow systems. The amount of oxygen that can be injected into the liquid (for a specific set of conditions) is approximately four times the amount that could be injected with an air system. In addition to process description, this chapter discusses covered and uncovered units, design considerations, design parameters and design procedure.

Key Words Pure oxygen • activated sludge • plug flow • complete mix • design parameters and procedure.

1. INTRODUCTION

As early as 1949, Okun (1) reported on research work being done for using pure oxygen as a substitute for air in the activated sludge process. The process was put into commercial use in 1970. The use of pure oxygen for activated sludge treatment has become competitive with the use of air owing to the development of efficient oxygen dissolution systems. The

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pure oxygen system is a high rate activated sludge system. The main advantages cited for the process include (2, 3):

(a) Reduced power requirements for dissolving oxygen in the wastewater
(b) Reduced aeration tank volume requirements
(c) Improved biokinetics of the activated sludge system
(d) The ability to treat high-strength wastewaters
(e) Decreased sludge volume, and
(f) Better settling sludge and decrease in bulking problems.

The pure oxygen system may be used for aeration in activated sludge systems that operate in either the plug flow or complete mix hydraulic regimes. It is readily adaptable to new or existing complete mix systems and can be used to upgrade and extend the life of overloaded plug-flow systems. The amount of oxygen that can be injected into the liquid (for a specific set of conditions) is approximately four times the amount that could be injected with an air system. Adjustment of pH may be necessary to maintain a proper balance between the CO_2 removed and buffer capacity of the wastewater (1, 4, 5).

2. PURE OXYGEN ACTIVATED SLUDGE, COVERED

2.1. Process Description

In the covered system, oxygenation is performed in a staged, covered reactor in which oxygen gas is recirculated within the system until it reaches a reduced level of purity and a decreased undissolved mass at which it can no longer be used and is vented to the atmosphere (6, 7). A flow diagram that illustrates this process is shown in Figure 7.1. High-purity oxygen gas (90% to 100% volume) enters the first stage of the system and flows concurrently with the wastewater being treated through the oxygenation basin. Pressure under the tank covers is essentially atmospheric, being held at 2 to 4 inches water column, sufficient to maintain oxygen gas feed control and prevent back-mixing from stage to stage. Effluent mixed liquor

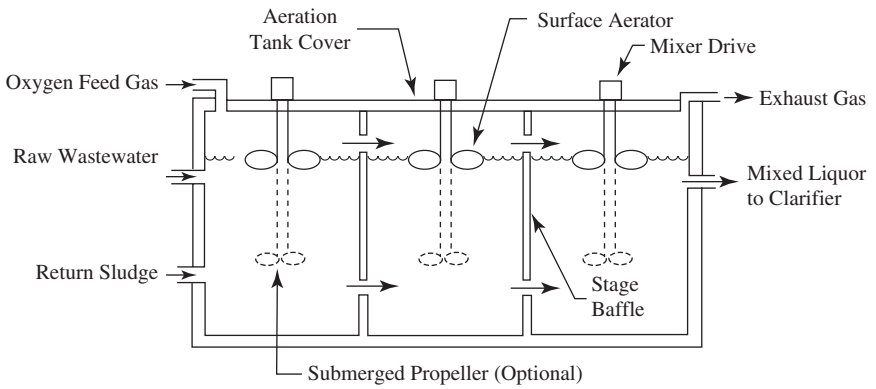

Fig. 7.1. Flow diagram for covered pure oxygen activated sludge process (2).

is separated in conventional gravity clarifiers, and the thickened sludge is recycled to the first stage for contact with influent wastewater.

Mass transfer and mixing within each stage are accomplished either with surface aerators or with a submerged-turbine rotating-sparge system. In the first case, mass transfer occurs in the gas space; in the latter, oxygen is sparged into the mixed liquor where mass transfer occurs from the oxygen bubbles to the bulk liquid. In both cases, the mass-transfer process is enhanced by the high oxygen-partial pressure maintained under the tank covers in each stage.

Volatile compounds are driven off to a certain extent in the oxygenation process and removed in the vent gas. Metals may also be expected to be partially removed, with accumulation in the sludge. High purity oxygen may be produced on-site by cryogenic or PSA (pressure swing adsorption) generators, or purchased as liquid oxygen produced off-site and stored at the treatment plant (7). Cost effectiveness of oxygen source depends upon plant size and process train.

Although flexibility is claimed to permit operation in any of the normally used flow regimes, i.e., plug flow, complete mix, step aeration, and contact stabilization, the method of oxygen contact employed favors the plug flow mode. Process may be designed to achieve: optimum carbonaceous oxidation only, combined carbonaceous and nitrogenous oxidation or optimum nitrogenous oxidation as a separate stage after secondary treatment. General concepts of biological treatment and evaluation of pure oxygen use in the activated sludge process can be found in References (8–13).

2.2. *Applications*

The most favorable situations for the application of covered pure oxygen activated sludge process include the following (2, 14, 15):

(a) Domestic and biologically degradable industrial wastewaters
(b) Upgrading existing air activated sludge plants
(c) New facilities to reduce construction cost whenever any of the following conditions are required:
    (1) Effective odor control
    (2) High effluent dissolved oxygen
    (3) Reduced quantity and higher concentration of waste sludge,
    (4) Reduced aeration detention time, and
    (5) More compact foot prints

One of the outstanding applications of the pure oxygen activated sludge process is the recently constructed 69th St. Advanced Wastewater Treatment Complex in The City of Houston, Texas (16). This plant serves one-third of the City's existing and projected population and doubles the City's sewage treatment capabilities. The plant's process is designed to meet the high quality effluent standards set by the Texas Department of Water Resources and to minimize noise, air and water pollution problems. The Complex consists of a 200-MGD (million gallons per day), two-stage, pure oxygen activated sludge advanced wastewater treatment plant and a 125 ton/day sludge processing and disposal plant. It represents a 400 percent expansion to the existing capacity of the Northside Treatment Facility, which was one of the first activated sludge plants in the United States (1916).

The wastewater exiting the primary treatment flows through the two stage activated sludge process where about 96 percent of BOD_5 is removed. The carbonaceous first stage is followed by a nitrification second stage. Pure oxygen from a 300 ton/day cryogenic oxygen generation plant is used in the activated sludge process. The pure oxygen system used in the reactor basins also alleviates odor and air pollution problems, while enhancing the appearance of the facility. The use of pure oxygen also permits greater loading rates per unit of site area. This choice of pure oxygen activated sludge process was the way in which a plant of this capacity could be accommodated onto the available project site (16).

2.3. *Design Criteria*

In the early applications of pure oxygen systems it was a common practice to design for MLSS concentrations in the high range of 3000 to 8000 mg/L and for short liquid detention times of 1 to 3 hours (17). However, recent research dealing with municipal wastewater treatment indicates that MLSS concentrations in the range of 1000 to 3000 mg/L constitute more suitable concentrations. SRTs in the neighborhood of 1 to 2 days are commonly used for treating municipal wastewater, although longer solids retention times are more prevalent when treating industrial wastewaters. Typical design criteria for carbonaceous BOD_5 oxidation are as follows (2, 18, 19):

(a)	Volumetric loading, lb BOD_5/d/1000 ft^3	100 to 200
(b)	F/M ratio, lb BOD_5/d/lb MLVSS	0.5 to 1.0
(c)	Oxygen requirement, lb O_2/lb COD removed	0.6 to 0.8
(d)	MLSS, mg/L	1000 to 6000
(e)	Aeration detention time, hr	1 to 3
(f)	Mixed liquor dissolved oxygen, mg/L	4 to 8
(g)	Oxygen required, lb O_2/lb BOD_5 removed	0.9 to 1.3

2.4. *Performance*

The usual production of CO_2 from biological reactions and its retention in the reactors tends to neutralize the buffering capacity of the flowing wastewater and results in increasing its acidity. pH values in the range of 6.0 to 6.5 or lower are common if no pH control is practiced (17, 20, 21). This pH reduction adversely impacts the nitrification process leading to the requirement of a longer SRT and a larger aeration tank. Consequently, the recommendation is to use two stage; a carbonaceous first stage followed by a nitrification second stage (20–24). Another option is to have staged nitrification-denitrification; the decreased pH in the nitrification process would be alleviated by the recovery of alkalinity through denitrification (21–25).

Modeling was used to control and optimize oxygen transfer in the pure oxygen activated sludge process. Simulations indicated that an optimal control system can reduce aerator power by 33% as compared to a conventional design, and reduce average oxygen feed gas by as much as 18% (26–28).

Typical performance data for the process are summarized in Table 7.1.

Table 7.1
Performance of covered pure oxygen activated sludge process

Type	Removal
Carbonaceous Oxidation	
COD removal, %	73–80
BOD_5 removal, %	89–95
Suspended solids removal, %	64–76
Nitrogenous Oxidation, NH_4-N removal in %	
Single stage with carbonaceous oxidation	20–90
Separate stage nitrification after carbonaceous oxidation	80–98
Residuals Generated, lb VSS/lb BOD_5 removed	0.42–1.0

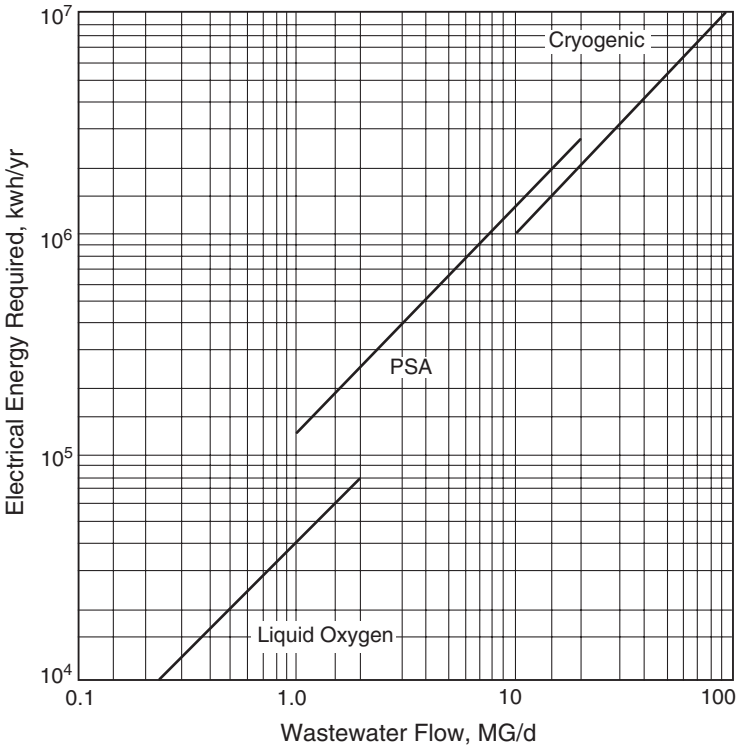

Fig. 7.2. Energy requirements for covered pure oxygen activated sludge process (2).

2.5. Energy Requirements

The covered pure oxygen activated sludge energy requirements are shown in Figure 7.2. The requirements are based on the following assumptions (2):

(a) Carbonaceous oxidation.
(b) Operating parameters: Oxygen requirement $= 1.2$ lb O_2/BOD_5 removed.

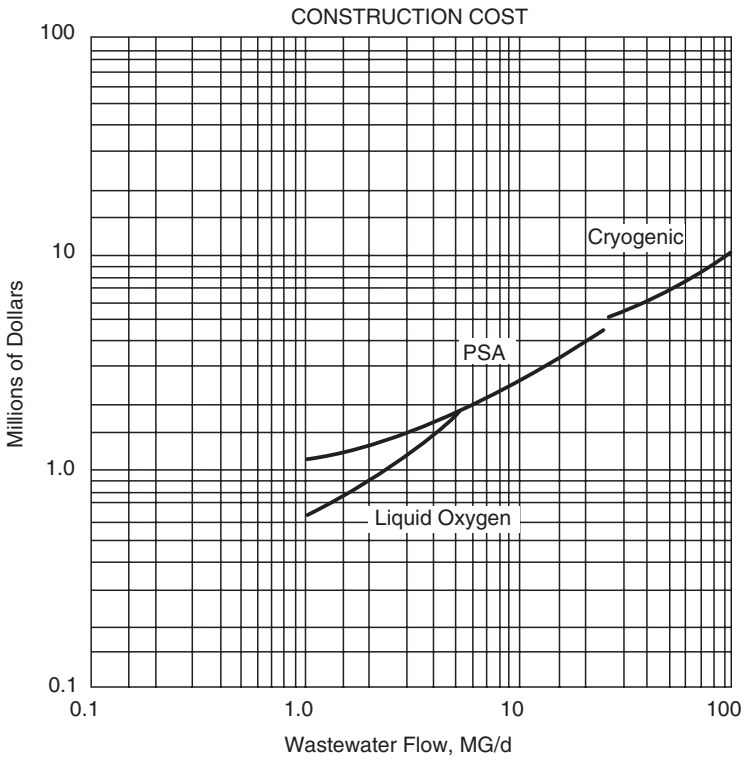

Fig. 7.3. Construction cost for covered pure oxygen activated sludge process (2).

(c) Water quality: Influent $BOD_5 = 130\,mg/L$; Effluent $BOD_5 = 20\,mg/L$.
(d) Oxygen transfer rate (OTR) includes oxygen production and oxygen dissolution.
(e) With cryogenic oxygen gas generation and surface aerators, OTR = 2.5 lb O_2/hp/h (wire to water) in wastewater.
(f) With pressure swing adsorption (PSA) oxygen gas generation and surface aerators, OTR = 2.0 lb O_2/hp/h (wire to water) in wastewater.
(g) Liquid O_2 supply and surface aerators, OTR = 6.5 lb O_2/hp/h (wire to water) in wastewater.

2.6. Costs

Construction and operation and maintenance costs (1979 Dollars, Utilities Index = 257.20) for covered pure oxygen activated sludge are shown in Figure 7.3 and Figure 7.4 respectively. To obtain the values in terms of the present 2008 U.S. Dollars, using the Cost Index for Utilities (Appendix A), multiply the costs by a factor of 552.16/257.20 = 2.15 (29). The costs are based on the following assumptions (2):

(a) Carbonaceous oxidation.
(b) Construction cost includes oxygenation basins, dissolutions equipment, oxygen generators and liquid oxygen feed/storage facilities, instrumentation (where applicable), and licensing fees.

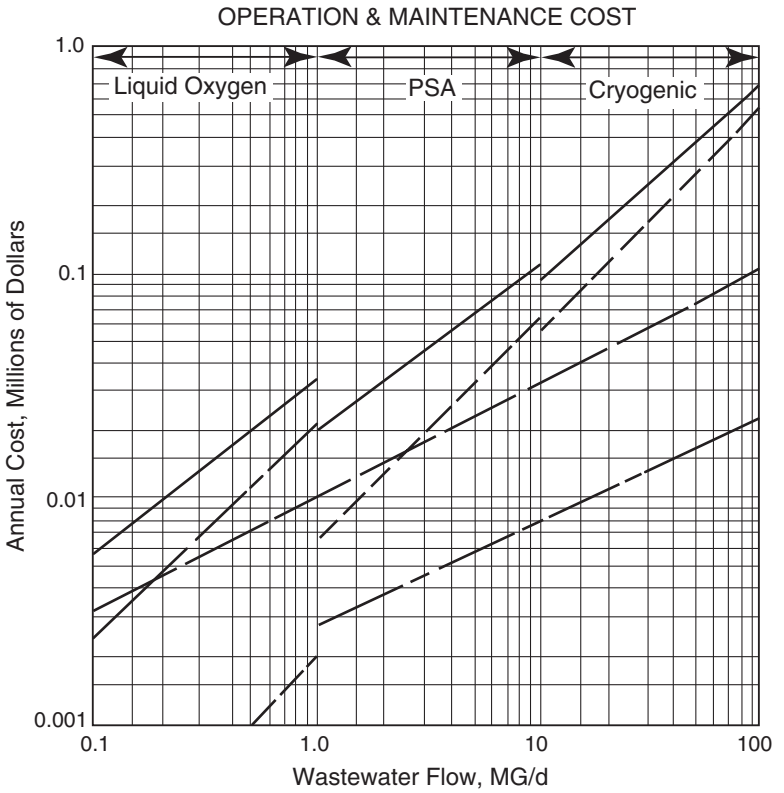

Fig. 7.4. Operation and maintenance cost for covered pure oxygen activated sludge process (2).

(c) Oxygen was assumed to be delivered as liquid oxygen for plants from 0.1 to 1 MGD size. For plants from 1.0 to 100 MGD, oxygen was assumed to be generated on-site.
(d) 1.2 lb O_2 supplied per lb BOD_5 removed.
(e) $MLVSS = 3100$ mg/L.
(f) $F/M = 0.5$ lb BOD_5/d/lb MLVSS.
(g) Volumetric loading = 97 lb BOD_5/d/1000 ft^3.
(h) Detention time = 2 hours (based on average daily flow).

3. PURE OXYGEN ACTIVATED SLUDGE, UNCOVERED

3.1. Description

The use of pure oxygen for activated sludge treatment has become competitive with the use of air owing to the development of efficient oxygen dissolution systems. The open tank oxygen system is a high rate activated sludge system. In the uncovered system, oxygenation is performed in an open reactor in which extremely fine porous diffusers are used to develop small oxygen gas bubbles that are completely dissolved before breaking surface in normal-depth tanks. The basic principles which apply in the transfer of oxygen in conventional

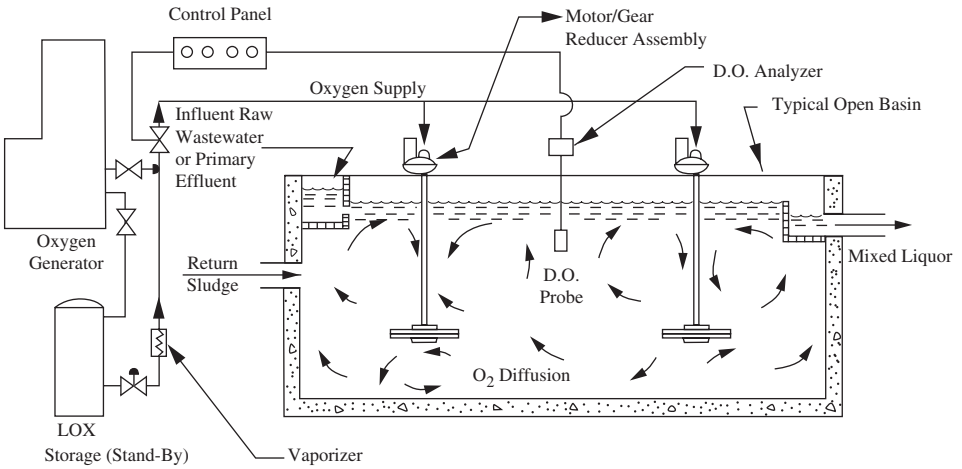

Fig. 7.5. Flow diagram for uncovered pure oxygen activated sludge process (2).

diffused air systems also apply to the open tank pure oxygen system (2). A flow diagram that illustrates this process is shown in Figure 7.5.

The pure oxygen open tank system produces ultra-fine bubbles with a correspondingly high gas surface area. These ultra-fine bubbles are of micron size, whereas "fine bubbles" normally produced in diffused air systems are in millimeter sizes. The complete oxygenation system is composed of an oxygen dissolution system comprised of:

(a) Rotating diffusers.
(b) A source of high-purity oxygen gas (normally, an on-site oxygen generator).
(c) Oxygen control system, which balances oxygen supply with oxygen demand through use of basin-located dissolved oxygen probes and control valves.

High purity oxygen may be produced on-site by cryogenic or PSA (Pressure Swing Adsorption) generators, or purchased as liquid oxygen produced off-site and stored at the treatment plant (7). Selection of cost effective oxygen source depends upon plant size and treatment train.

The influent to the system enters the oxygenation tank and is mixed with return activated sludge. The mixed liquor is continuously and thoroughly mixed using low energy mechanical agitation deep in the mixed liquor. Mixing is produced by radial turbine impellers located on both surfaces (top and bottom) of the rotating diffusion discs. Pure oxygen gas in the form of micron-size bubbles is simultaneously introduced into the tank to accomplish mass oxygen transfer. The rotating diffuser is a gear-driven disc-shaped diffusion device equipped with a porous medium to assist in the diffusion process. As the diffuser rotates at constant speed in the mixed liquor, hydraulic shear wipes bubbles from the medium before they have an opportunity to coalesce and enlarge.

Operation in any of the normally used flow regimes, i.e., plug flow, complete mix, step aeration, and contact stabilization, can be used as conditions dictate because the method of oxygen contact employed does not favor one particular operating mode. System may be designed to optimize carbonaceous (BOD_5) oxidation, combined carbonaceous (BOD_5) and nitrogenous (NOD) oxidation as a single stage, or nitrogenous oxidation as a separate stage after secondary treatment.

3.2. Applications

The most favorable situations for the application of uncovered pure oxygen activated sludge process include the following:

(a) Domestic and biologically degradable industrial wastewaters.
(b) Plant flows greater than 1 MGD.
(c) Upgrading existing air activated sludge plants.
(d) New facilities to reduce construction cost whenever any of the following conditions are required:
  (1) High effluent dissolved oxygen.
  (2) Reduced quantity and higher concentration of waste sludge.
  (3) Reduced aeration detention time.

3.3. Design Criteria

The reported design criteria for uncovered systems are as follows (2):

(a)	Volumetric Loading, lb BOD_5/d/1000 ft^3	100 to 200
(b)	*F/M* ratio, lb BOD_5/d/lb MLVSS	0.5 to 1.0
(c)	Oxygen requirement,	
	i. lb O_2/lb BOD_5 removed	0.9 to 1.3
	ii. lb O_2/lb COD removed	0.6 to 0.8
(d)	Aeration detention time, hr	1 to 3 (based on avg. daily flow)
(e)	Mixed liquor dissolved oxygen, mg/L	2 to 6
(f)	MLSS, mg/L	3000 to 6000

3.4. Performance

Removal efficiencies of various pollutants are similar to those of activated sludge and vary with mode of operation, aeration detention time, and character of influent wastewater. Examples of operational and pilot test data are shown in Table 7.2.

3.5. Energy Requirements

The uncovered pure oxygen activated sludge energy requirements are shown in Figure 7.6. The requirements are based on the following assumptions (2):

(a) Carbonaceous oxidation.
(b) Operating parameters: Oxygen requirement = 1.2 lb O_2/BOD_5 removed.
(c) Water quality: Influent BOD_5 = 130 mg/L; Effluent BOD_5 = 20 mg/L.
(d) Oxygen transfer rate (OTR) includes oxygen production and oxygen dissolution.

Table 7.2
Performance of uncovered pure oxygen activated sludge process

Type	Removal
Carbonaceous Oxidation	
COD removal, %	60–85
BOD_5 removal, %	75–95
Suspended solids removal, %	60–90
Nitrogenous Oxidation, NH_4-N removal in %	
Single stage with carbonaceous oxidation	20–90
Separate stage nitrification after carbonaceous oxidation	80–98
Residuals Generated, lb VSS/lb BOD_5 removed	0.42–1.0

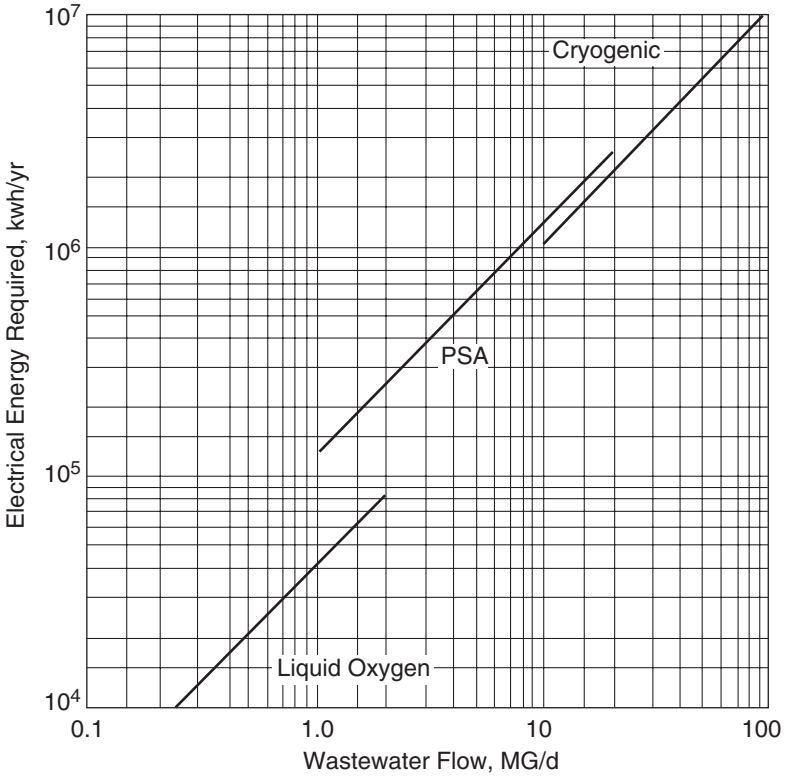

Fig. 7.6. Energy requirements for uncovered pure oxygen activated sludge process (2).

(e) With cryogenic oxygen gas generation and surface aerators, OTR = 2.5 lb O_2/hp/h (wire to water) in wastewater.
(f) With pressure swing adsorption (PSA) oxygen gas generation and surface aerators, OTR = 2.0 lb O_2/hp/h (wire to water) in wastewater.
(g) Liquid O_2 supply and surface aerators, OTR = 6.5 lb O_2/hp/hr (wire to water) in wastewater.

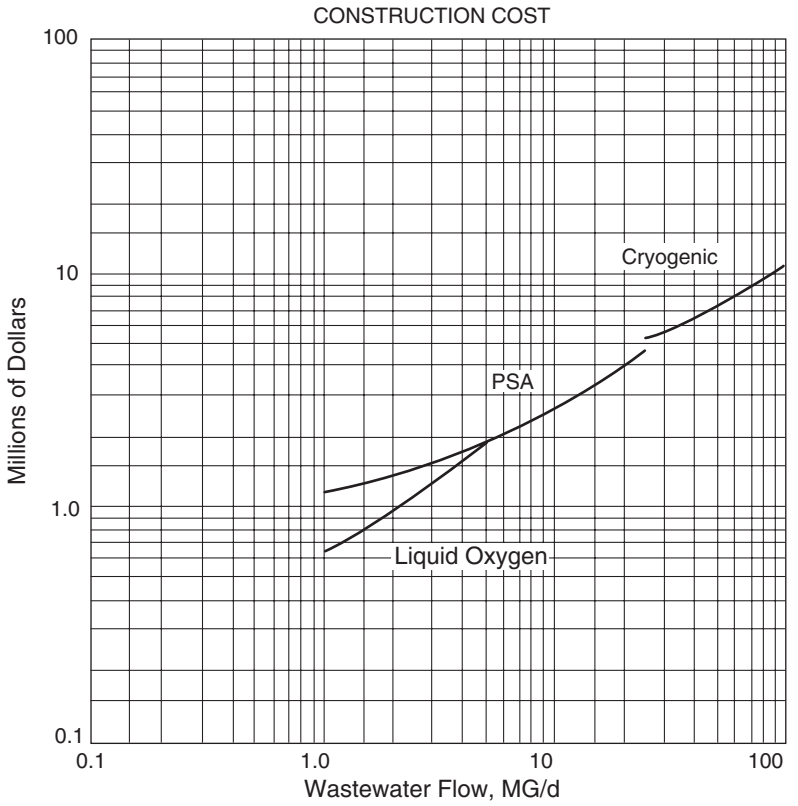

Fig. 7.7. Construction cost for uncovered pure oxygen activated sludge process (2).

3.6. Costs

Construction and operation and maintenance costs (1979 Dollars, Utilities Index = 257.20) for uncovered pure oxygen activated sludge are shown in Figures 7.7 and 7.8 respectively. To obtain the values in terms of the present 2008 U.S. Dollars, using the Cost Index for Utilities (Appendix A), multiply the costs by a factor of $552.16/257.20 = 2.15$ (29). The costs are based on the following assumptions (2):

(a) Carbonaceous oxidation.
(b) Construction cost includes oxygenation basins, dissolutions equipment, oxygen generators and liquid oxygen feed/storage facilities, instrumentation (where applicable), and licensing fees.
(c) Oxygen was assumed to be delivered as liquid oxygen for plants from 0.1 to 1 MGD size. For plants from 1.0 to 100 MGD, oxygen was assumed to be generated on-site.
(d) 1.2 lb O_2 supplied per lb BOD_5 removed.
(e) $MLVSS = 3{,}100$ mg/L.
(f) $F/M = 0.5$ lb BOD_5/d/lb MLVSS.
(g) Volumetric loading = 97 lb BOD_5/d/1000 ft^3.
(h) Detention time = 2 hours (based on average daily flow).

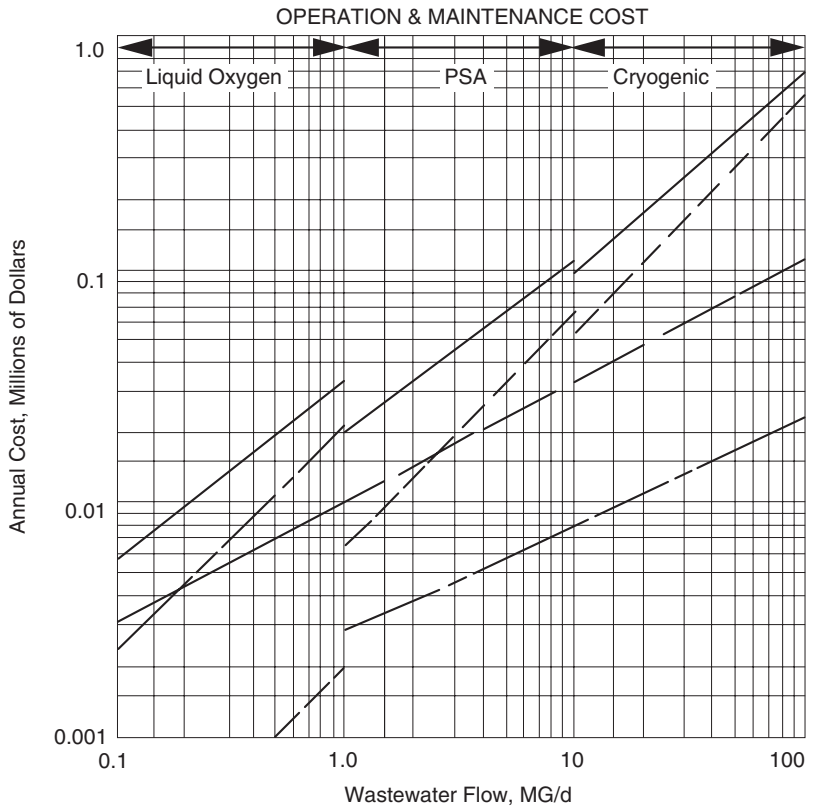

Fig. 7.8. Operation and maintenance cost for uncovered pure oxygen activated sludge process (2).

4. DESIGN CONSIDERATIONS

4.1. Input Data

The input data required for a design problem includes the following (30):

(a) Wastewater flows (average and peak). In case of high variability, a statistical distribution should be provided.
(b) Wastewater strength.
  (1) BOD_5 (soluble and total), mg/L.
  (2) COD and/or TOC (maximum and minimum), mg/L.
  (3) Suspended solids, mg/L.
  (4) Volatile suspended solids (VSS), mg/L.
  (5) Nonbiodegradable fraction of VSS, mg/L.
(c) Other characteristics.
  (1) pH.
  (2) Acidity and/or alkalinity, mg/L.
  (3) Nitrogen, mg/L.
  (4) Phosphorus (total and soluble), mg/L.

(5) Oils and greases, mg/L.
(6) Heavy metals, mg/L.
(7) Toxic or special characteristics (e.g., phenols), mg/L.
(8) Temperature, °F or °C.

(d) Effluent quality requirements.
(1) BOD_5, mg/L.
(2) SS, mg/L.
(3) TKN, mg/L.
(4) P, mg/L.
(5) Total nitrogen (TKN + NO_3-N), mg/L.
(6) Settleable solids, mg/L.

4.2. Design Parameters

Design parameters which are either to be known or to be assumed include the following (11, 12, 30, 32):

(a) Reaction rate constants and coefficients.
McKinney
$K_m = 15/h$ at 20°C
$K_s = 10.4/h$ at 20°C
$K_e = 0.02/h$ at 20°C
Eckenfelder
$k = 0.0007 - 0.002\,L/mg/h$
$a = 0.73$
$a' = 0.52$
$b = 0.075/d$
$b' = 0.15/d$
$f = 0.40$
$f' = 0.53$
(b) $F/M = 0.25$ to 1.0
(c) Volumetric loading = 150–200
(d) t = 2–4 h
(e) t_s = 3–20 d, depending on application
(f) MLSS = 4000–7000 mg/L, mean 5000 mg/L
(g) MLVSS = 3200–5600 mg/L
(h) Q_r/Q = 0.25–0.50 = recycle ratio
(i) lb O_2/lb BOD_r = 1.0–1.5
(j) lb solids/lb BOD_r = 0.30–0.45
(k) θ = 1.0–1.03
(l) Efficiency = > 90%

4.3. Design Procedure

The following is a summarized presentation of the design procedure for a pure oxygen activated sludge process. For further details, the reader is referred to the extensive literature dealing with this subject (10–13, 17–19, 30–37.)

4.3.1. McKinney's Approach

(1) Assume the following design parameters.
   (a) Metabolism constant (K_m).
   (b) Synthesis factor (K_s).
   (c) Endogenous respiration factor (K_e).
   (d) Temperature correction coefficient (θ).
   (e) Hydraulic detention time (t).
   (f) Solids retention time (t_s).

(2) Adjust metabolism constant, synthesis factor, and endogenous respiration factor for temperature

$$k_T = k_{20}\theta^{(T-20)} \tag{1}$$

where
k_T = rate constant at desired temperature T, °C
k_{20} = rate constant at 20°C
θ = temperature coefficient
T = temperature, °C

(3) Determine size of the aeration tank

$$V = Q_{avg}\, t/24 \tag{2}$$

where
V = volume of tank, MG
Q_{avg} = average flow, MGD
t = hydraulic detention time, h

(4) Determine soluble effluent BOD_5

$$F_e = F_i/1 + K_m t \tag{3}$$

where
F_e = effluent BOD_5, mg/L
F_i = influent BOD_5, mg/L
K_m = metabolism constant 1/h (15/h at 20°C)
t = hydraulic detention time, h
and check $F_e < 10$ mg/L; if $F_e > 10$ mg/L, increase t and recalculate new F_e

(5) Calculate the MLSS concentration

$$M_T = M_a + M_e + M_i + M_{ii} \tag{4}$$
$$M_a = K_s F_e/(K_e + t_s/24) \tag{5}$$
$$M_e = 0.2 K_e M_a t_s(24) \tag{6}$$
$$M_i = SS_i \times (24 t_s/t) \tag{7}$$
$$M_{ii} = SS_i \times (24 t_s/t) + 0.1(M_a + M_e) \tag{8}$$

where
M_T = total mass, mg/L
M_a = living, active mass, mg/L
M_e = endogenous mass, mg/L

M_i = inert nonbiodegradable organic mass, mg/L
M_{ii} = inert inorganic suspended solids, mg/L
K_s = synthesis factor, 1/h (10.4/h at 20°C)
F_e = effluent BOD_5, mg/L
K_e = endogenous respiration factor, 1/h (0.02/h at 20°C)
T_s = solids retention time, *d*
SS_i = inert organic *SS* in influent, mg/L
= *VSS* × percent nonbiodegradable (≈ 0.4 *VSS* for municipal waste)
SS_{ii} = inert inorganic *SS* fraction in the influent
and check M_T against 4000–7000 mg/L; vary t_s or t until M_T falls within desired range.

(6) Check organic loading against 0.25–1.0

$$F/M = 24F_i/M_T t \tag{9}$$

where
F/M = food-to-microorganism ratio
F_i = influent BOD_5, mg/L
M_T = total mass, mg/L
t = hydraulic detention time, h
If $F/M <$ lower limit, it is possible to reduce t and recalculate M_T
If $F/M >$ upper limit, increase t and recalculate M_T

(7) Calculate the oxygen requirements.

(a) Select the oxygen uptake rate. The average rate of oxygen demand, if the waste load is uniform, is given by

$$dO/dt = [1.5(F_i - F_e)/t] - [1.42(M_a + M_e)/24t_s] \tag{10}$$

where
dO/dt = average oxygen uptake rate under uniform flow conditions, mg/L
F_i = influent BOD_5, mg/L
F_e = soluble effluent BOD_5, mg/L
t = hydraulic detention time, h
M_a = living, active mass, mg/L
M_e = endogenous mass, mg/L
t_s = solids retention time, *d*
Under conditions where the load varies, the oxygen uptake is equal to the synthesis oxygen demand plus the endogenous respiration oxygen demand or

$$dO/dt = [0.5(F_i - F_e)/t](Q_p/Q_{avg}) + 1.14K_e M_a \tag{11}$$

$$\text{lb } O_2/\text{h} = dO/dt \times V \times 8.34 \tag{12}$$

where
Q_p = peak flow, MGD
Q_{avg} = average flow, MGD
K_e = endogenous respiration factor, 1/h (0.02/h at 20°C)

(b) Check oxygen supplied per pound of BOD removed > 1.25

$$\text{lb } O_2/\text{h} \times 24/Q(F_i - F_e) \times 8.34 \tag{13}$$

(8) Design aeration system

Check horsepower for complete mixing ≥ 0.1 hp/1000 gal; select the larger horsepower

(a) Diffused aeration system

1. Assume the following design parameters.
   a. Standard transfer efficiency, percent, from manufacturer.
   b. O_2 transfer in waste/O_2 transfer in water ≈ 0.9.
   c. O_2 saturation in waste/O_2 saturation in water ≈ 0.9.
   d. Correction factor for pressure ≈ 1.0.
2. Select summer temperature (25 to 30°C) and determine (from standard tables) O_2 saturation.
3. Adjust standard transfer efficiency to operating conditions.

$$OTE = STE\ \{[(C_s)_T^{\beta p} - C_L]/9.17\}\alpha(1.02)^{T-20} \tag{14}$$

where
OTE = operating transfer efficiency, %
STE = standard transfer efficiency, %
$(C_s)_T = O_2$ saturation at selected summer temperature T, °C, mg/L
$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
p = correction factor for pressure ≈ 1.0
C_L = minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
T = temperature, °C

4. Calculate required air flow.

$$R_a = O_2(10^5)(7.48)/(OTE)(1440\ \text{min/d})(V)(CF) \tag{15}$$

where
R_a = required air flow, cfm/1000 ft^3
O_2 = oxygen required, lb/d
OTE = operating transfer efficiency, %
V = volume of basin, gal
CF = correction factor, lb O_2/ft^3 air

(b) Mechanical aeration system.

1. Assume the following design parameters.
   a. Standard transfer efficiency, lb/hp/h (0 dissolved oxygen, 20°C and tap water).
   b. O_2 transfer in waste/O_2 transfer in water ≈ 0.9.
   c. O_2 saturation in waste/O_2 saturation in water ≈ 0.9.
   d. Correction factor for pressure ≈ 1.0.
2. Select summer temperature (25 to 30°C), and determine (from standard tables) O_2 saturation.
3. Adjust standard transfer efficiency to operating conditions

$$OTE = STE\ \{[(C_s)_T^{\beta p} - C_L]/9.17\}\alpha(1.02)^{T-20} \tag{16}$$

where
OTE = operating transfer efficiency, lb O_2/hp/h
STE = standard transfer efficiency, lb O_2/hp/h
$(C_s)_T = O_2$ saturation at selected summer temperature T, °C, mg/L

$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
p = correction factor for pressure ≈ 1.0
C_L = minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
T = temperature, °C

4. Calculate horsepower requirement

$$\text{hp} = O_2 \times 1000/(\text{OTE})(\text{lb } O_2/\text{hp/h})(24)(V) \tag{17}$$

where
hp = horsepower required/1000 gal
O_2 = oxygen required, lb/d
OTE = operating transfer efficiency, lb O_2/hp/h
V = volume of basin, gal

(9) Calculate sludge production and determine pounds of sludge wasted per day

$$\Delta M_T = 8.34\, M_T V/t_s \tag{18}$$

where
ΔM_T = sludge produced, lb/d
M_T = total mass concentration, mg/L
V = volume of aeration tank, MG
t_s = solids retention time, d

(10) Check solids produced per pound of BOD removed.

$$\text{lb solids/lb BOD}_5 = \Delta M_T/8.34Q(F_i - F_e) \tag{19}$$

where
ΔM_T = sludge produced, lb/d
Q = flow, MGD
F_i = influent BOD_5, mg/L
F_e = effluent BOD_5, mg/L

(11) Calculate sludge recycle ratio.

$$Q_r/Q = M_T/(M_u - M_T) \tag{20}$$

where
Q_r = volume of recycled sludge, MGD
Q = flow, MGD
M_T = total mass concentration, mg/L
M_u = solids concentration in return sludge, mg/L

(12) Calculate total effluent BOD_5.

$$(\text{BOD}_5)_{\text{eff}} = S_e + 0.84(SS_{\text{eff}})(M_a/M_T)(0.76) \tag{21}$$

where
S_e = effluent soluble BOD, mg/L
SS_{eff} = effluent suspended solids, mg/L
M_a = living mass, mg/L
M_T = total mass, mg/L

(13) Determine nutrient requirements for nitrogen and phosphorous.

$$\mathrm{N} = 0.123\Delta M_{\mathrm{T}} \text{ (or } \Delta X_{\mathrm{V}}) \tag{22}$$

$$\mathrm{P} = 0.026\Delta M_{\mathrm{T}} \text{ (or } \Delta X_{\mathrm{V}}) \tag{23}$$

where
ΔM_{T} = sludge produced, lb/d
ΔX_{V} = sludge produced, lb/d
and check against BOD:N:P = 100:5:1

4.3.2. *Eckenfelder's Approach*

(1) Assume the following design parameters when unknown.
(a) BOD removal rate constant (k).
(b) Fraction of BOD synthesized (a).
(c) Fraction of BOD oxidized for energy (a').
(d) Endogenous respiration rate (b and b').
(e) Mixed liquor suspended solids (MLSS).
(f) Mixed liquor volatile suspended solids (MLVSS).
(g) Food-to-microorganism ratio (F/M).
(h) Nonbiodegradable fraction of VSS in influent (f).
(i) Degradable fraction of the MLVSS (f').
(j) Temperature correction coefficient (θ).

(2) Adjust rate constant for temperature

$$k_{\mathrm{T}} = k_{20}\theta^{(T-20)} \tag{24}$$

where
k_{T} = rate constant at desired temperature T, °C
k_{20} = rate constant at 20°C
θ = temperature coefficient
T = temperature, °C

(3) Determine the size of the aeration tank by first determining the detention time t.

$$t = 24\ \mathrm{S_o}/[(X_{\mathrm{V}})(F/M)] \tag{25}$$

where
t = detention time, h
S_{o} = influent BOD, mg/L
X_{V} = MLVSS, mg/L
F/M = food-to-microorganism ratio

(4) Check detention time for treatability.

$$S_{\mathrm{e}}/S_{\mathrm{o}} = 1/(1 + kX_V t) \tag{26}$$

where
S_{e} = $\mathrm{BOD_5}$ (soluble) in effluent, mg/L
S_{o} = $\mathrm{BOD_5}$ in influent, mg/L
k = BOD removal rate constant, L/mg/h
X_{V} = MLVSS, mg/L
t = detention time, h
Solve for t and compare with t above and select the larger.

(5) Calculate the volume of aeration tank.

$$V = Q_{\text{avg}}(t/24) \tag{27}$$

where
V = volume, MG
Q_{avg} = average daily flow, MGD
t = detention time, h

(6) Calculate oxygen requirements.

$$\text{dO}/\text{d}t = (a' S_{\text{r}}/t) + b' X_{\text{V}} \tag{28}$$

or

$$\text{O}_2 = a'(S_{\text{r}})(Q_{\text{avg}})(8.34) + b'(X_{\text{V}})(V)(8.34) \tag{29}$$

where
$\text{dO}/\text{d}t$ = oxygen uptake rate, mg/L/h
a' = fraction of BOD oxidized for energy
S_{r} = BOD removed ($S_{\text{o}} - S_{\text{e}}$), mg/L
t = detention time, h
b' = endogenous respiration rate, 1/d
X_{V} = MLVSS, mg/l
O_2 = oxygen requirement, lb/d
Q_{avg} = average flow rate, MGD
V = volume of aeration tank, MG
and check the oxygen supplied against ≥ 1.25

$$\text{lb O}_2/\text{lb BOD}_{\text{r}} = \text{O}_2/8.34\, QS_{\text{r}} \tag{30}$$

where
O_2 = oxygen required, lb/d
Q = flow, MGD
S_{r} = BOD removed, mg/L

(7) Design aeration system and check horsepower supply for mixing against horsepower required for complete mixing ≤ 0.1 hp/1000 gal.
   (a) Diffused aeration system.
      1. Assume the following design parameters
         a. Standard transfer efficiency, percent, from manufacturer.
         b. O_2 transfer in waste/O_2 transfer in water ≈ 0.9.
         c. O_2 saturation in waste/O_2 saturation in water ≈ 0.9.
         d. Correction factor for pressure ≈ 1.0.
      2. Select summer temperature (25 to 30°C) and determine (from standard tables) O_2 saturation.
      3. Adjust standard transfer efficiency to operating conditions.

$$OTE = STE\,\{[(C_{\text{s}})_{\text{T}}^{\beta\text{p}} - C_{\text{L}}]/9.17\}\alpha(1.02)^{\text{T}-20} \tag{31}$$

where
OTE = operating transfer efficiency, %
STE = standard transfer efficiency, %

$(C_s)_T = O_2$ saturation at selected summer temperature T, °C, mg/L
$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
$p =$ correction factor for pressure ≈ 1.0
$C_L =$ minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
$T =$ temperature, °C

4. Calculate required air flow.

$$R_a = O_2(10^5)(7.43)/(OTE)(1440 \text{ min/d})V(CF) \tag{32}$$

where
$R_a =$ required air flow, cfm/1,000 ft^3
$O_2 =$ required oxygen, lb/d
$OTE =$ operating transfer efficiency, %
$V =$ volume of basin, gal
$CF =$ correction factor, lb O_2/ft^3 air

(b) Mechanical aeration system.

1. Assume the following design parameters
   a. Standard transfer efficiency, lb/hp/h (0 dissolved oxygen, 20°C, and tap water).
   b. O_2 transfer in waste/O_2 transfer in water ≈ 0.9.
   c. O_2 saturation in waste/O_2 saturation in water ≈ 0.9.
   d. Correction factor for pressure ≈ 1.0.
2. Select summer temperature (25 to 30°C) and determine (from standard tables) O_2 saturation.
3. Adjust standard transfer efficiency to operating conditions.

$$OTE = STE\ \{[(C_s)_T^{\beta\ p} - C_L]/9.17\}\alpha(1.02)^{T-20} \tag{33}$$

where
$OTE =$ operating transfer efficiency, lb O_2/hp/h
$STE =$ standard transfer efficiency, lb O_2/hp/h
$(C_s)_T = O_2$ saturation at selected summer temperature T, °C, mg/L
$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
$p =$ correction factor for pressure ≈ 1.0
$C_L =$ minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
$T =$ temperature, °C

4. Calculate horsepower requirement.

$$\text{hp} = O_2 \times 1000/(OTE)(\text{lb } O_2/\text{hp/h})(24)(\text{V}) \tag{34}$$

where
hp = horsepower required/1000 gal
$O_2 =$ oxygen required, lb/d
$OTE =$ operating transfer efficiency, lb O_2/hp/h
$V =$ volume of basin, gal

(8) Calculate sludge production.

$$\Delta X_V = [aS_rQ - bX_VV + fQ(VSS) + Q(SS - VSS)]8.34 \tag{35}$$

where
ΔX_V = sludge produced, lb/d
a = fraction of BOD removed synthesized to cell material
S_r = BOD removed, mg/L
Q = average flow, MGD
b = endogenous respiration rate, 1/d
X_V = volatile solids in raw sludge, mg/L
V = volume of basin, MG
f = nonbiodegradable fraction of influent VSS
VSS = volatile suspended solids in effluent, mg/L
SS = suspended solids in influent, mg/L

(9) Check ΔX_V against > 0.30–0.45.

$$\text{lb solids}/(\text{lb BOD}_r) = XV/S_r(Q)(8.34) \tag{36}$$

where
ΔX_V = sludge produced, lb/d
S_r = BOD removed, mg/L
Q = flow, MGD

(10) Calculate sludge recycle ratio.

$$Q_r/Q = X_a/(X_u - X_a) \tag{37}$$

where
Q_r/Q = sludge recycle ratio
Q_r = volume of recycled sludge, MGD
Q = average flow, MGD
X_a = MLSS, mg/L
X_u = solids concentration in return sludge, mg/L

(11) Calculate solids retention time.

$$SRT = (V)X_a(8.34)/\Delta X_a \tag{38}$$

where
SRT = solids retention time, d
V = Volume of aeration tank, MG
X_a = MLSS, mg/L
$\Delta X_a = \Delta X_V/\%$ volatile
ΔX_V = volatile sludge produced, mg/L

(12) Calculate effluent BOD_5.

$$(BOD_5)_{eff} = S_e + 0.84\ (X_V)_{eff} f' \tag{39}$$

where
S_e = effluent soluble BOD, mg/L
$(X_V)_{eff}$ = effluent volatile suspended solids, mg/L
f' = degradable fraction of MLVSS

(13) Determine nutrient requirements for nitrogen and phosphorus.

$$\mathrm{N} = 0.123\ \Delta M_{\mathrm{T}} (\text{or } \Delta X_{\mathrm{V}}) \tag{40}$$

$$\mathrm{P} = 0.026\ \Delta M_{\mathrm{T}} (\text{or } \Delta X_{\mathrm{V}}) \tag{41}$$

where
ΔM_{T} = sludge produced, lb/d
ΔX_{V} = sludge produced, lb/d
and check against BOD : N : P = 100 : 5 : 1

4.4. Output Data

(a) Aeration tank.
1. Reaction rate constant, L/mg/h.
2. Sludge produced per BOD removed.
3. Endogenous respiration rate (b, b').
4. O_2 = used per BOD removed.
5. Influent nonbiodegradable VSS (f).
6. Effluent degradable VSS (f').
7. lb BOD/lb MLVSS/d (F/M ratio).
8. Mixed liquor SS, mg/L (MLSS).
9. Mixed liquor VSS, mg/L(MLVSS).
10. Aeration time, h.
11. Volume of aeration tank, MG.
12. Oxygen required, lb/d.
13. Sludge produced, lb/d.
14. Nitrogen requirement, lb/d.
15. Phosphorus requirement, lb/d.
16. Sludge recycle ratio, %.
17. Solids retention time, d.

(b) Diffused Aeration System.
1. Standard transfer efficiency, %.
2. Operating transfer efficiency, %.
3. Required air flow, cfm/1000 ft^3.

(c) Mechanical aeration system.
1. Standard transfer efficiency, lb O_2/hp/h.
2. Operating transfer efficiency, lb O_2/hp/h.
3. Horsepower required.

5. DESIGN EXAMPLE

Assume that you have the following design parameters for a pure oxygen activated sludge process:

$k = 0.0012$ L/mg/h
$a = 0.73$
$a' = 0.5$
$b = 0.075$/d, $b' = 0.15$/d

$MLSS = X_a = 3000\,\text{mg/L}$
$MLVSS = X_V = 2400\,\text{mg/L}$
$F/M = 0.40$ lb BOD/lb *MLVSS*/d
$f = 0.40$
$f' = 0.53$
$\theta = 1.03$

Determine the following:

(a) Adjust the BOD removal rate constant for winter conditions, $T = 15°C$.
(b) Determine the size of the aeration tank by first determining the detention time t.
(c) Check detention time for treatability.
(d) Calculate the volume of aeration tank.
(e) Calculate oxygen requirements.
(f) Design aeration system (mechanical surface).
(g) Calculate sludge production.
(h) Calculate sludge recycle ratio.
(i) Calculate solids retention time.
(j) Calculate effluent BOD_5.
(k) Determine nutrient requirements for nitrogen and phosphorus.

Solution

(a) Adjust the BOD removal rate constant for winter conditions, $T = 15°C$.

$$k_T = k_{20}\theta^{(T-20)}$$

where
k_T = rate constant at desired temperature T, °C
k_{20} = rate constant at 20°C, 0.0012 L/mg/h
θ = temperature coefficient, 1.03
T = temperature, 15°C

$$k_T = 0.0012\ (1.03)^{15-20}$$
$$k_T = 0.0010\ \text{L/mg/h}$$

(b) Determine the size of the aeration tank by first determining the detention time t.

$$t = 24S_o/[(X_V)(F/M)]$$

where
t = detention time, h
S_o = influent BOD, 200 mg/L
X_V = *MLVSS*, 2400 mg/L
F/M = food-to-microorganism ratio, 0.40 lb BOD/lb MLVSS/d

$$t = 24(200)/2400(0.40)$$
$$t = 5.0\,\text{h}$$

(c) Check detention time for treatability.

$$S_e/S_o = 1/(1 + kX_V t)$$

where
S_e = BOD_5 (soluble) in effluent, 10 mg/L
S_o = BOD_5 in influent, 200 mg/L
k = BOD removal rate constant, 0.001 L/mg/h
X_V = *MLVSS*, mg/L
t = detention time, h
Solve for t and compare with t above and select the larger.

$$10/200 = 1/1 + 0.001 \times 2400t$$
$$\text{t} = 7.9\,\text{h} > 5.0\,\text{h}$$

(d) Calculate the volume of aeration tank.

$$V = Q_{avg}(t/24)$$

where
V = volume, MG
Q_{avg} = average daily flow, 1.0 MGD
t = detention time, 7.9 h

$$V = 1.0(7.9/24)$$
$$V = 0.329\,\text{MG}$$

(e) Calculate oxygen requirements.

$$O_2 = a'(S_r)(Q_{avg})(8.34) + b'(X_V)(V)(8.34)$$

where
a' = fraction of BOD oxidized for energy, 0.52
S_r = BOD removed ($S_o - S_e$), 190 mg/L
b' = endogenous respiration rate, 0.15/d
X_V = *MLVSS*, 2400 mg/l
O_2 = oxygen requirement, lb/d
Q_{avg} = average flow rate, 1.0 MGD
V = volume of aeration tank, 0.329 MG

$$O_2 = 0.52(190)1.0(8.34) + 0.15(2{,}400)0.329(8.34)$$
$$O_2 = 1{,}812\,\text{lb/d}$$

and check the oxygen supplied against ≥ 1.25

$$\text{lb } O_2/\text{lb } BOD_r = O_2/8.34QS_r$$

where
O_2 = oxygen required, 1,812 lb/d
Q = flow, 1.0 MGD
S_r = BOD removed, 190 mg/L

$$\text{lb } O_2/\text{lb BOD}_r = 1{,}812/(8.34)1.0(190)$$

$$\text{lb } O_2/\text{lb BOD}_r = 1.14 < 1.25$$

Therefore

$O_2 = 1.25\,\text{BOD}_r$
$O_2 = 1.25\,Q\,S_r(8.34)$
$O_2 = 1.25\ (1.0)\ 190\ (8.34)$
$O_2 = 1{,}980\,\text{lb/d}$

(f) Design aeration system (mechanical surface).

1. Assume the following design parameters
   a. Standard transfer efficiency, *STE* 5%.
   b. O_2 transfer in waste/O_2 transfer in water $\alpha \approx 0.9$.
   c. O_2 saturation in waste/O_2 saturation in water $\beta \approx 0.9$.
   d. Correction factor for pressure $p \approx 1.0$.
2. Select summer temperature (25°C to 30°C) and determine from tables O_2 saturation.

$$T = 25°\text{C},$$

$$(C_s)_T = 8.2\,\text{mg/L}$$

3. Adjust standard transfer efficiency to operating conditions.

$$OTE = STE\,\{[(C_s)_T^{\beta\ p} - C_L]/9.17\}\alpha(1.02)^{T-20}$$

where
OTE = operating transfer efficiency, %
STE = standard transfer efficiency, 5%
$(C_s)_T = O_2$ saturation at selected summer temperature T, °C, 8.2 mg/L
$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
p = correction factor for pressure ≈ 1.0
C_L = minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
T = temperature, 25°C

$$OTE = 5.0\{[(8.2(0.9)1.0 - 2.0]/9.17\}0.9(1.02)^{25-20}$$

$$OTE = 2.9\%\ \text{or lb } O_2/\text{hp/h}$$

4. Calculate horsepower requirement.

$$\text{hp} = O_2 \times 1000/(OTE)(\text{lb } O_2/\text{hp/h})(24)(V)$$

where
hp = horsepower required/1000 gal
O_2 = oxygen required, 1980 lb/d
OTE = operating transfer efficiency, 2.9% or lb O_2/hp/h
V = volume of basin, gal (0.329 MG = 329,000 gal)

$$\text{hp} = 1{,}980/2.9(24)329$$

$$\text{hp} = 0.09\,\text{hp}/1{,}000\,\text{gal} < 0.1$$

Therefore, use hp = 0.1 hp/1000 gal
hp = 0.1 (V)1000
hp = 0.1 (0.329) 1000
hp = 32.9 hp (use 35)

(g) Calculate sludge production

$$\Delta X_{\mathrm{V}} = [aS_r Q - bX_V V + fQ(VSS) + Q(SS - VSS)]8.34$$

where
ΔX_{V} = sludge produced, lb/d
a = fraction of BOD removed synthesized to cell material, 0.73
S_{r} = BOD removed, 190 mg/L
Q = average flow, 1.0 MGD
b = endogenous respiration rate, 0.075/d
X_{V} = volatile solids in raw sludge, 2400 mg/L
V = volume of basin, 0.329 MG
f = nonbiodegradable fraction of influent *VSS*, 0.40
VSS = volatile suspended solids in effluent, 150 mg/L
SS = suspended solids in influent, 200 mg/L

$$\Delta X_{\mathrm{V}} = [0.73(190)1.0 - 0.075(2400)0.329 + 1.0(150)0.40 + 1.0(200 - 150)]/8.34$$

$$\Delta X_{\mathrm{V}} = 1580\,\mathrm{lb/d}$$

Check ΔX_{V} against > 0.30–0.45

$$\text{lb solids}/(\text{lb BOD}_{\mathrm{r}}) = XV/S_{\mathrm{r}}(Q)(8.34)$$

where
ΔX_{V} = sludge produced, 1580 lb/d
S_{r} = BOD removed, 190 mg/L
Q = average flow, 1.0 MGD

$$\text{lb solids/lb BOD}_{\mathrm{r}} = 1{,}580/190(1.0)8.34$$
$$= 1.0 > 0.3 \text{ to } 0.45(\mathrm{OK})$$

(h) Calculate sludge recycle ratio.

$$Q_r/Q = X_a/(X_u - X_a)$$

where
Q_r/Q = sludge recycle ratio
Q_r = volume of recycled sludge, MGD
Q = average flow, 1.0 MGD
X_a = MLSS, 3,000 mg/L
X_u = solids concentration in return sludge, 10,000 mg/L

$$Q_r/1.0 = 3000/10{,}000 - 3000$$
$$Q_r = 0.429\,\mathrm{MGD}$$

(i) Calculate solids retention time.

$$SRT = (V)X_a(8.34)/\Delta \mathrm{x_a}$$

where
SRT = solids retention time, d
V = Volume of aeration tank, 0.329 MG
X_a = MLSS, 3000 mg/L
ΔX_V = volatile sludge produced, mg/L

$$\Delta X_a = \Delta \mathrm{X_V}/\%\ \text{volatile} = 1{,}580/0.80 = 1{,}975\ \mathrm{lb/d}$$
$$SRT = 0.329(3,000)8.34/1975$$
$$SRT = 4.2\ \mathrm{d}$$

(j) Calculate effluent BOD_5.

$$(\mathrm{BOD_5})_{\mathrm{eff}} = S_\mathrm{e} + 0.84(\mathrm{x_V})_{\mathrm{eff}}\ f'$$

where
S_e = effluent soluble BOD, 10 mg/L
$(X_\mathrm{V})_{\mathrm{eff}}$ = effluent volatile suspended solids, 20 mg/L
f' = degradable fraction of MLVSS, 0.53

$$(\mathrm{BOD_5})_{\mathrm{eff}} = 10 + 0.84(20)0.53$$
$$(\mathrm{BOD_5})_{\mathrm{eff}} = 19\ \mathrm{mg/L}$$

(k) Determine nutrient requirements for nitrogen and phosphorus.

$$\mathrm{N} = 0.123\ \Delta M_\mathrm{T}\ (\mathrm{or}\Delta X_\mathrm{V})$$
$$\mathrm{P} = 0.026\ \Delta M_\mathrm{T}\ (\mathrm{or}\Delta X_\mathrm{V})$$

where
ΔM_T = sludge produced, lb/d
ΔX_V = volatile sludge produced, lb/d

$$\mathrm{N} = 0.123(1,580)$$
$$\mathrm{N} = 194\ \mathrm{lb/d}$$
$$\mathrm{P} = 0.026(1,580)$$
$$\mathrm{P} = 41\ \mathrm{lb/d}$$

and check against $\mathrm{BOD:N:P} = 100:5:1$

$$\text{N in influent} = 30\ \mathrm{mg/L(Q)}8.34$$
$$= 30(1.0)8.34$$
$$= 250\ \mathrm{lb/d} > 194\ \text{required}$$

Therefore, no additional N is needed

$$\text{P in influent} = 15\,\text{mg/L}(1.0)8.34$$
$$= 125\,\text{lb/d} > 41\,\text{lb/d required}$$

Therefore, no additional P is needed

NOMENCLATURE

a = fraction of BOD removed synthesized to cell material
a' = fraction of BOD oxidized for energy
b = endogenous respiration rate, 1/d
b' = endogenous respiration rate, 1/d
CF = correction factor, lb O_2/ft^3 air
C_L = minimum dissolved oxygen to be maintained in the basin ≥ 2.0 mg/L
$(C_s)_T$ = O_2 saturation at selected summer temperature T, °C, mg/L
dO/dt = average oxygen uptake rate under uniform flow conditions, mg/L
f = nonbiodegradable fraction of influent VSS
f' = degradable fraction of MLVSS
F/M = food-to-microorganism ratio
F_e = effluent BOD_5, mg/L
F_i = influent BOD_5, mg/L
hp = horsepower required/1000 gal
k = BOD removal rate constant, L/mg/h
k_T = rate constant at desired temperature T, °C
k_{20} = rate constant at 20°C
K_e = endogenous respiration factor, 1/hr (0.02/hr at 20°C)
K_m = metabolism constant 1/hr (15/hr at 20°C)
K_s = synthesis factor, 1/hr (10.4/hr at 20°C)
M_a = living, active mass, mg/L
M_e = endogenous mass, mg/L
M_i = inert nonbiodegradable organic mass, mg/L
M_{ii} = inert inorganic suspended solids, mg/L
M_T = total mass, mg/L
M_u = solids concentration in return sludge, mg/L
MLSS = Mixed liquor suspended solids
MLVSS = Mixed liquor volatile suspended solids
O_2 = oxygen required, lb/d
OTE = operating transfer efficiency, %
p = correction factor for pressure ≈ 1.0
Q = flow, MGD
Q_{avg} = average flow, MGD
Q_p = peak flow, MGD
Q_r = volume of recycled sludge, MGD

Q_r/Q = sludge recycle ratio
R_a = required air flow, cfm/1,000 ft^3
Se = effluent soluble BOD, mg/L
S_o = influent BOD, mg/L
S_r = BOD removed ($S_o - S_e$), mg/L
SRT = solids retention time, d
SS = suspended solids in influent, mg/L
SS_{eff} = effluent suspended solids, mg/L
x = inert organic SS in influent, mg/L
STE = standard transfer efficiency, %
t = hydraulic detention time, hr
t_s = solids retention time, d
T = temperature, °C
T_s = solids retention time, d
V = volume of tank, MG
X = MLSS, mg/L
X_u = solids concentration in return sludge, mg/L
X_V = MLVSS, mg/L
$(X_V)_{eff}$ = effluent volatile suspended solids, mg/L
$\alpha = O_2$ transfer in waste/O_2 transfer in water ≈ 0.9
$\beta = O_2$ saturation in waste/O_2 saturation in water ≈ 0.9
θ = temperature coefficient
$\Delta X_a = \Delta X_V$/% volatile
$\Delta M_T = \Delta X_V$ = sludge produced, lb/d

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APPENDIX

US Army Corps of Engineers Civil Works Construction Yearly Average cost Index for Utilities (29)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

* US ACE. Yearly average Cost Index for Utilities. In: *Civil Works Construction Cost Index System Manual*, 110-2-1304, U.S. Army Corps of Engineers, Washington, DC, PP 44. PDF file is available on the Internet at http://www.nww.usace.army.mil/cost) (2007).

8
Waste Stabilization Ponds and Lagoons

Nazih K. Shammas, Lawrence K. Wang, and Zucheng Wu

CONTENTS

Abstract One of the simplest forms of biological treatment processes is the stabilization pond or stabilization lagoon. It is also the most common industrial wastewater treatment facility. This versatile installation serves many basic purposes, including: (a) storage or impoundment of wastewater; (b) settling and removal of suspended solids; (c) storage or impoundment of settled solids; (d) equalization; (e) aeration; (f) biological treatment; and (g) evaporation. The relative simplicity and low operating costs of a stabilization pond make it the preferred technology for handling, treatment and disposal of industrial wastewater as well as municipal wastewater for small communities. Besides the description of ponds complex ecological system and the complicated reactions that take place, the chapter covers the system variables, design criteria, process control, capital and operating costs, applications and examples of process design.

Key Words Stabilization pond • lagoon • wastewater treatment • process design • process control and costs.

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1. CONCEPTS AND PHYSICAL BEHAVIOR

One of the simplest forms of biological treatment processes is the stabilization pond or stabilization lagoon. It is also the most common industrial wastewater treatment facility. This versatile installation serves many basic purposes, including: (a) storage or impoundment of wastewater; (b) settling and removal of suspended solids; (c) storage or impoundment of settled solids; (d) equalization; (e) aeration; (f) biological treatment; and (g) evaporation. The relative simplicity and low operating costs of a stabilization pond make it the preferred technology for handling, treatment and disposal of industrial wastewater as well as municipal wastewater for small communities. The main advantages and disadvantages of stabilization pond are listed below (1):

(a) Advantages

1. Low operational and maintenance cost
2. Lagoons provide effective treatment with minimal threat to the environment
3. Work well in clay soils where conventional subsurface on-site absorption fields will not work.

(b) Disadvantages

1. Lagoons must be constructed in clay soil or be lined to prevent leakage
2. May overflow occasionally during extended periods of heavy rainfall
3. If there are extended periods of overcast windless days, offensive odors may occur for a brief time
4. Lagoons usually recover rapidly if this occurs
5. Can not be installed on a small lot. Takes up a relatively large space
6. Lagoons are not aesthetically acceptable to some people. Some people consider lagoons unsightly and unsafe.
7. As with any other open body of water, there is some potential danger. Although lagoons are required to be fenced, this does not always prevent access by people or pets.

Based on a survey by the US EPA (2), treatment systems in the general category of "stabilization ponds" usually serve small communities of whom 90% have populations of 10,000 persons or fewer.

1.1. Pond Ecology and Process Reactions

The simplicity of a stabilization pond belies the fact that complicated reactions – both chemically and biologically – often occur in a pond. A highly complex ecological system exists that is subject to diurnal and seasonal variations. Figure 8.1 depicts the major reactions taking place in a stabilization pond. Eight major processes can be identified in active stabilization ponds. They are

1. sedimentation
2. aerobic decomposition
3. anaerobic fermentation
4. bacterial – algal symbiosis
5. oxygen transfer across the water surface
6. Sulfur bacteria actions
7. evaporation, and
8. seepage

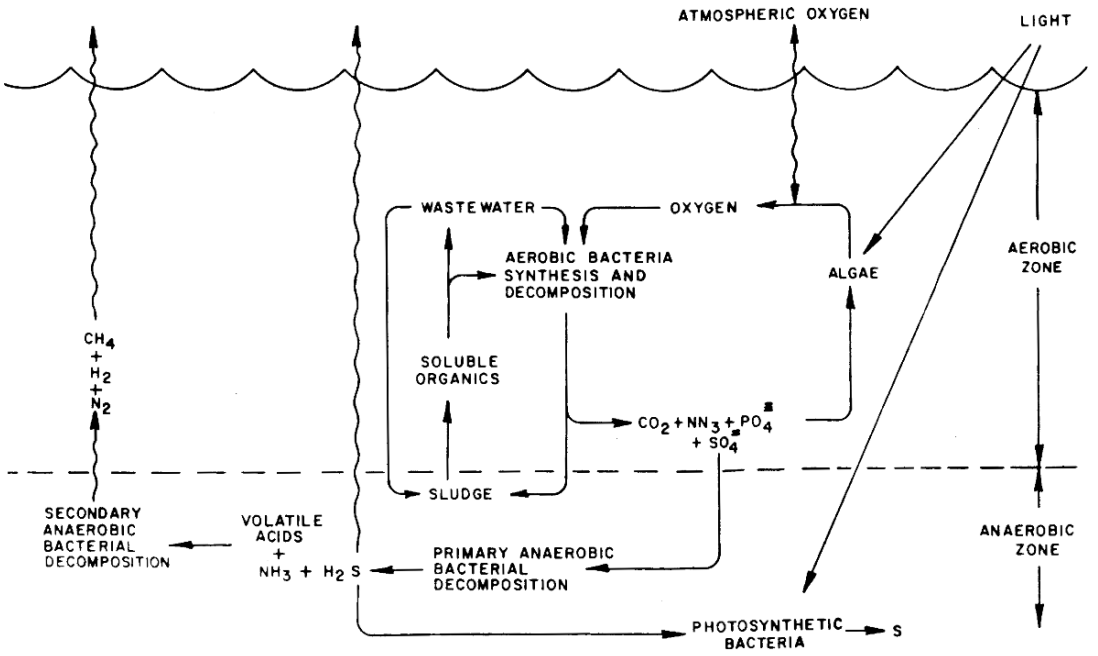

Fig. 8.1. Ecology and major reactions of stabilization ponds.

Depending on the type of stabilization pond, some or all of these reactions can take place simultaneously. Each major reaction will be discussed below, while the types of stabilization ponds will be discussed in detail later.

1.1.1. Sedimentation

Suspended matter from incoming wastewater and fecal matter from worms and insects will precipitate. Such precipitation can be enhanced significantly by chemical and biological flocculation in the pond. Vigorous photosynthetic actions by algal cells result in a rise of pH, promoting the formation of calcium and magnesium flocs in the alkaline condition. From 80 to 90% of suspended matter can be precipitated in a few hours, depending on the temperature, hydraulic flow regime, and depth of a pond. The bioflocculation of synthesized bacterial cells and algal cells also constitute a part of the sedimentation regime.

1.1.2. Aerobic Decomposition

As long as the dissolved oxygen is maintained above the critical level of 0.1 to 0.2 mg/L, aerobic oxidation of biodegradable organics will take place in the pond similar to that in an activated sludge process. While nitrification can take place in an activated sludge process, it seldom occurs in a stabilization pond because ammonia is readily used by algae for growth. At high pH, nitrogen stripping or precipitation as $MgNH_4PO_4$ could also take place before nitrification can become established.

1.1.3. Anaerobic Fermentation

Anaerobic decomposition of organic matter takes place in an established anaerobic zone in a pond. Both acid and alkaline fermentation occur concurrently to yield gases H_2S, CH_4, H_2, N_2, etc. that may escape the pond. Odor production is often accompanied by a vigorous fermentation at high temperature and in the presence of high organic loading rates.

1.1.4. Bacterial-Algal Symbiosis

Where turbidity of the pond water is low and sunlight is plentiful, a bacterial – algal symbiosis often occurs. In the upper zone of the pond, algae grow in abundance and through photosynthetic action, could supersaturate the zone with molecular oxygen. The abundant supply of oxygen supports the active aerobic oxidation by bacteria that in turn yield the inorganic nutrients NH_3, PO_4^{3-}, and CO_2 in particular, to meet the demand in algal growth.

Photosynthetic activity in a stabilization pond demands large amount of CO_2 for algae synthesis. There are three CO_2 sources upon which algae synthesis can draw for use: (a) CO_2 as an end product of bacterial oxidation and fermentation of organics in the pond; (b) CO_2 from the atmosphere; and (c) CO_2 from the inorganic carbon species in the $CO_2-HCO_3^--CO_3^{2-}$ system. As Figure 8.2 shows, a vigorous photosynthetic activity may lead to rapid depletion of CO_2 and a consequent rise of pH in the pond. This indeed was observed in lagoons as well as in natural water systems by Day (3), Rich (4), Goldman (5), Kerr (6), and King (7), and so on. Strong bacterial activity is therefore required to support the photosynthesis when other environmental conditions such as light intensity and temperature are favorable. Oxygen is produced through photosynthesis in direct proportion to the algal growth. If one assumes an empirical algal cell composition of $C_7H_{8.1}O_{2.5}N$, the following overall reaction can take place:

$$7.6\,CO_2 + 17.7\,H_2O + HN_4^+ \xrightarrow{\text{light}} C_7H_{8.1}O_{2.5}N + 7.6\,O_2 + 15.2\,H_2O + H^+ - 886\,\text{kcal} \quad (1)$$

It can be seen from Equation 1 that 2.3 g of CO_2 are required for every gram of algal cell synthesized. Respectively, the O_2 production is 1.67 g. This production in many cases can be an adequate supply of oxygen to satisfy the bacterial oxidation process without any other source of supply.

1.1.5. Oxygen Transfer across the Water Surface

Oxygen saturation in pond water can occur during the day when photosynthesis is at its maximum. Consequently, O_2 can escape the pond. More often than not, O_2 is below saturation level and atmospheric pressure reaeration takes place. The rate of O_2 transfer depends on the deficit (or surplus if oversaturation) as well as on surface renewal. The rate constant K_2 can be expressed in the following form:

$$K_2 = \frac{-1}{t} \log \frac{C_s - C}{C_s - C_0} \quad (2)$$

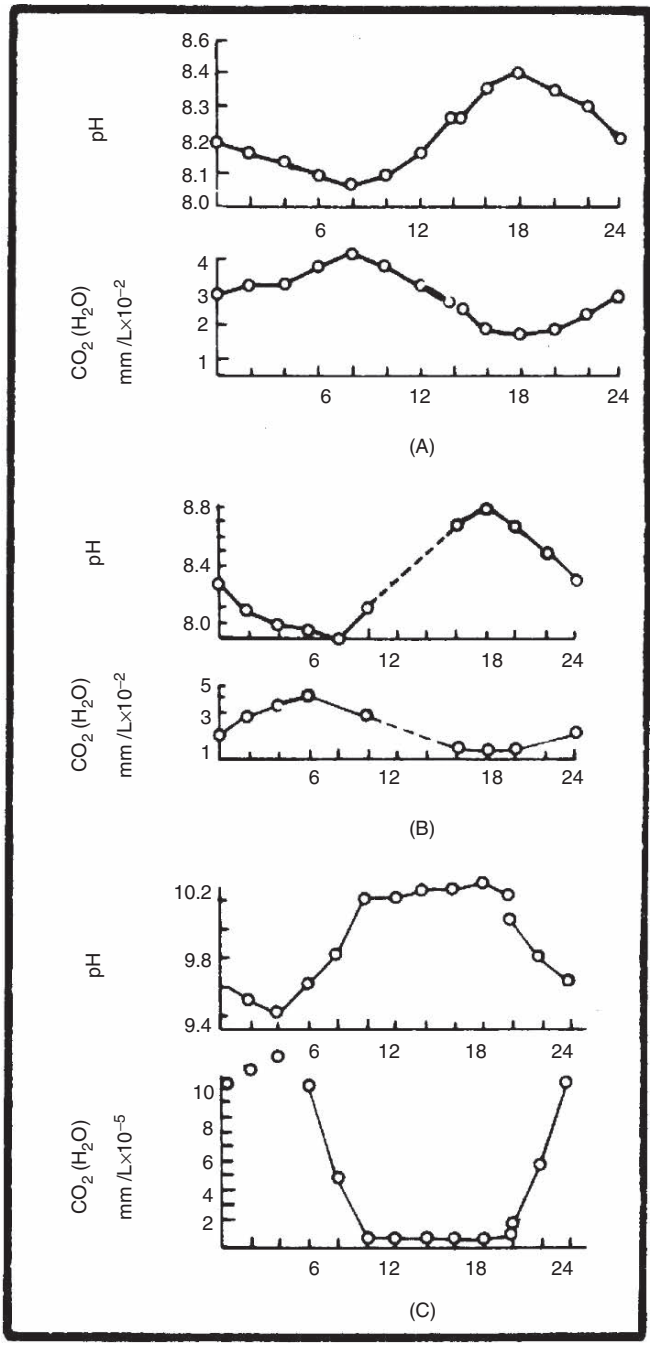

Fig. 8.2. Diurnal pH and CO_2 patterns observed in lagoons.

where t = time, C_s = saturation concentration, C = concentration at any time, C_0 = initial O_2 concentration, and K_2 = rate constant. When $C_s > C > C_0$, K_2 is a reaeration constant. When $C_s < C < C_0$, K_2 is a deaeration constant (oxygen stripping).

It is difficult to evaluate the rate constant K, because it is a function of surface renewal and temperature. Methods to evaluate K values for streams do not apply because only wind-induced mixing is important for oxygen transfer across the pond water surface. Ekman has developed empirical formulations for wind-induced current velocity and depth of circulation as follows (8):

$$P = \frac{V_0}{W} = 0.0127\sqrt{\sin\theta} \tag{3}$$

$$\frac{D}{W} = 11.1\sqrt{\sin\theta} \tag{4}$$

where V_0 = surface-current velocity caused by the wind and W wind speed; while (V_0/W) is a term called proportional surface velocity of the water (with observed magnitudes of p lies between 1×10^{-2} and 5×10^{-2} a function of latitude: D circulation depth or depth of frictional resistance.

Knowing both p and D, the following equations according to Fair (9) can be used for estimation of the reaeration constant K_2;

$$G^2 = 93 \times 10^{-3}(pW)^3/(\mu g F) \tag{5}$$

$$K_2 = 29G^3/D \tag{6}$$

where W is wind speed in miles/hr and F is the fetch in miles, μ is the absolute viscosity with both μ and g in ft-lb-s units, G is the mean temporal velocity gradient in s^{-1} and D is the circulation depth in feet.

Banks (10) expressed the oxygen transfer coefficient K_L as a function of wind speed as follows:

$$K_L = kU^m \tag{7}$$

in which $k = 4.19 \times 10^{-6}$; 1.8×10^{-6}; 0.32×10^{-6} and m = 1/2, 1, and 2 for U = small, medium, and large wind speed, respectively. K_L and U are both expressed in m/s. Solution to Equation (7) is shown in Figure 8.3. After a value of K_L is obtained, the reaeration constant K_2, can be calculated as:

$$K_2 = K_L/D \tag{8}$$

1.1.6. Sulfur Bacteria Actions

As depicted in Figure 8.1, aerobic oxidation of sulfur compounds in wastewater produces sulfate, whereas anaerobic bacteria reduce sulfate to sulfide. The most characteristic specie of the colorless group of bacteria reducing sulfides is *Desulfovibrio sp.* whose optimal growth occurs at pH 7.0 and ORP (oxidation reduction potential) at −100 to −300 mV. It has been

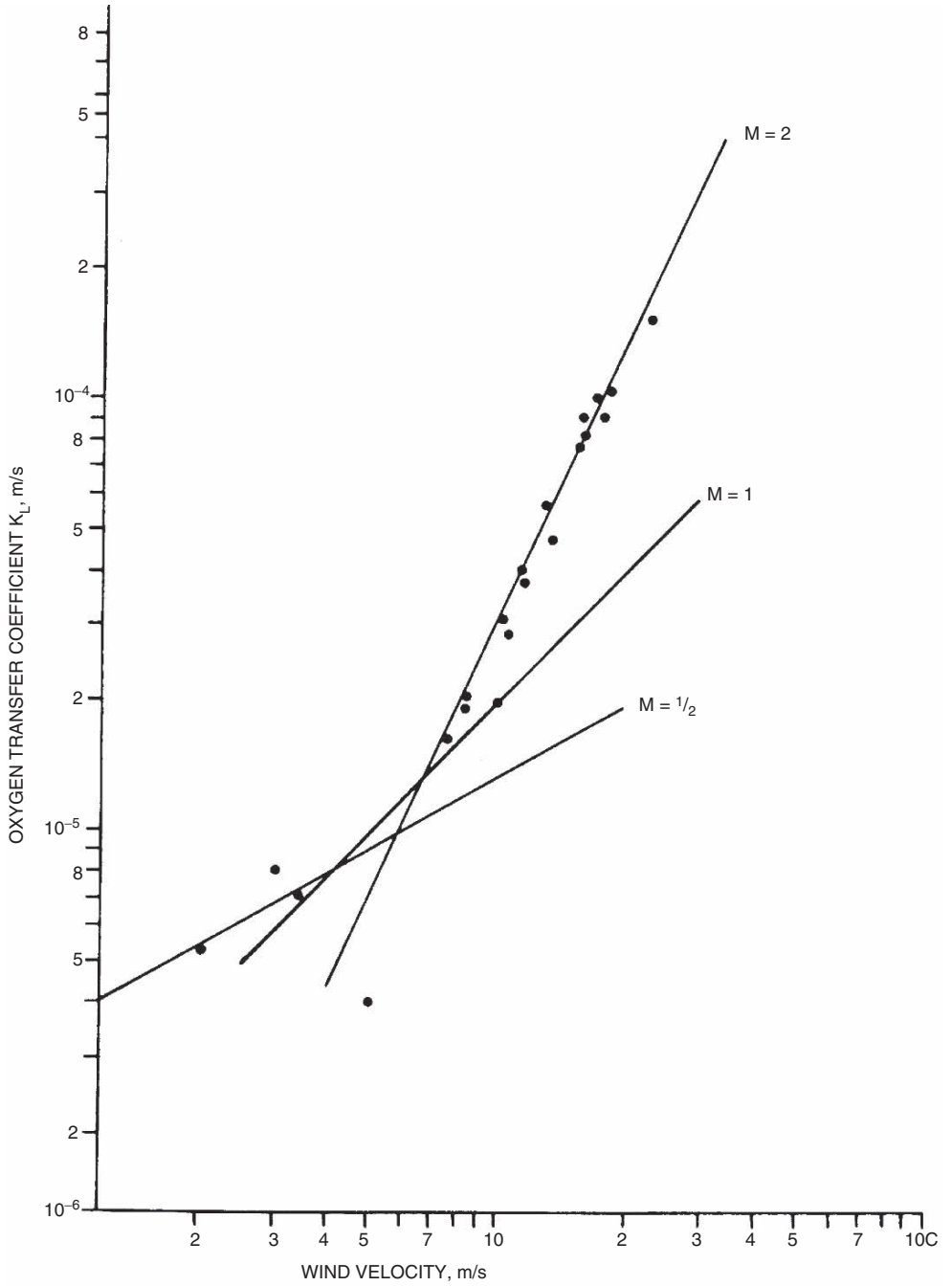

Fig. 8.3. Transfer coefficients as functions of various powers of wind velocity.

reported (11) that this colorless group of bacteria does not grow at potentials higher than +27 mV.

Sulfides can be oxidized using either molecular oxygen or CO_2 as the hydrogen acceptor. Because only the top layer of the pond is conducive to the growth of colorless sulfur bacteria oxidizing sulfides, they are not often found in stabilization ponds. The group of microorganisms using CO_2 as the hydrogen acceptor is the photosynthetic sulfur bacteria. Included in this group are the green *Chlorobacteriaceae* and the purple *Thirohodaceae*. The oxidation of sulfide is carried out in two steps in the absence of oxygen:

$$CO_2 + 2H_2S \rightarrow (CH_2O) + H_2O + 2S \tag{9}$$

$$3CO_2 + 2S + 5H_2O \rightarrow (CH_2O) + 4H^+ + 2SO_4^{2-} \tag{10}$$

It has been found by Gloyna (12) that algae virtually disappear when the sulfide concentration reaches 6 to 7 mg/L in the pond water. Daytime sulfide concentration is 5% to 25% less than the dark periods because of sulfide oxidation by the photosynthetic oxygen and by the photosynthetic sulfur bacteria. Increases in sulfide concentration from near zero to near 1 and 2 mg/L reduce the ORP 80 and 120 mV respectively.

1.1.7. Evaporation

Evaporation loss of pond water is normally insignificant except in warm and dry climates. It is not an unusual practice, however, to have lagoons for total storage of wastewater providing no outlet and depending completely on evaporation loss. Rates of evaporation from ponds vary with temperature, vapor pressures of the water and the air in contact with it, wind speed, barometric pressure, and the salt content of water.

Relating evaporation to many of the factors previously mentioned; Rohwer (13) developed a formula in the following form:

$$E = 0.497(1 - 1.32 \times 10^{-2} P_a)(1 + 0.268W)(V - v) \tag{11}$$

in which E = evaporation in in/d; P_a = barometric pressure in inches of mercury; W is wind speed in miles/hr; and V and v are vapor pressures in inches of mercury at the water temperature and dewpoint temperature of the atmosphere, respectively. Wind and temperature affect evaporation profoundly. At the temperatures of natural waters, the vapor pressure is almost doubled for every rise of 10°C. Wind stimulates evaporation by displacing moisture-laden films with relatively dry air.

1.1.8. Seepage

Depending on the wastewater quality and soil characteristics, the water loss from seepage can be substantial. The seepage flow should be stopped if the quality of a nearby ground water supply is threatened.

Hart (14) found substantial water losses in anaerobic ponds constructed on sandy soils. In ponds 7 ft deep, the infiltration rate ranged from 6.6 cm/d at an early stage of the operation to 4.1 cm/d 2 years later. Davis (15) reported stabilization ponds treating dairy wastes having an infiltration rate of 122 cm/d in the beginning. The rate dropped to 0.5 cm/d after 4 months of operation.

Natural sealing of stabilization ponds takes place gradually. According to Chang (16), the sealing is attributable to the physical entrapment of organic particulates in the pores of soil followed by the growth of slime-forming microorganisms. Evidence is also presented by Matthew (17) that the sodium adsorption ratio of the wastewater can influence the permeability of stabilization pond soils. The sodium adsorption ratio is defined by US Salinity Laboratory as follows:

$$\text{SAR(sodium adsorption ratio)} = [Na^+]/\{0.5[Ca^{2+}] + 0.5[Mg^{2+}]\}^{0.5} \quad (12)$$

The normal correlation is one of decreasing permeability as the SAR is increasing.

Sealing of ponds can be done artificially by injecting clay, bentonite, or asphalt, or by placing liners over the bottom and sides of a pond. The latter procedure has become a commonly accepted practice because it is simple and effective. More discussion of pond sealing will be presented later in the design section.

1.2. Biology of Stabilization Ponds

The biological population depends on the type of stabilization pond, the influent wastewater characteristics, the time of the day, and the season of the year. Because of limited control in operation, variations of biological population in stabilization ponds are much more apparent than in any other biological process.

1.2.1. Bacterial Population

Unlike activated sludge processes, stabilization ponds are operated at a very low bacterial population. One seldom finds a bacterial concentration higher than 100 mg/L in stabilization ponds. Total plate counts of bacteria generally fall in the range of 10^6 to 10^7/mL.

Coli-aerogenes group is predominant in stabilization ponds. If carbohydrates are limited in the wastewater because of pretreatment or decomposition in the sewer line before reaching the pond, proteolytic and lipolytic bacteria will predominate. Gann (18) reported that *Achromobacter*, *Pseudomonas*, and *Flavobacterium* are dominant in an aerobic pond, accounting for 90% of the total bacterial count. Parker (19) also reported abundant number of *Achromobacter* and *Bacillus* (lipolytic) and other proteolytic groups in anaerobic lagoons. Gram-positive bacteria, including *Streptococcus faecalis* and *Bacillus*, commonly occur in aerobic stabilization ponds, although they are never present in significant numbers.

Bacteria die-off in ponds has been reported by Pratt (20), Vladimirava (21), Davis (22), and others. Although individual algal species exert little influence on the dieoff of enteric bacteria, mixtures of axenic algal cultures increase the dieoff rates. *Escherichia*, *Pseudomonas*, and *Serratia* exhibit aftergrowth potential whereas *Proteus*, *Alcaligenes*, *Enterobacter*, *Salmonella*, *Shigella*, and *Vibrio* do not. The die-off rates of these bacteria are increased by anaerobic pretreatment of the wastewater.

Desulfovibrio sp. are abundant in stabilization ponds when substantial amounts of sulfate are present in wastewater. The green *Chlorobacteriacea* and the purple *Thirohodaceae* then oxidize H_2S to sulfur elements and sulfates in the presence of light. Both can use short infrared radiation according to Hutchinson (23) and Rabinowitch (24). High H_2S concentration can promote an overgrowth of the purple sulfur bacteria to the extent that the entire pond is covered

Table 8.1
Physiology of photosynthetic plants and bacteria

Description	Green plants	Green bacteria	Purple bacteria
Source of reducing power	H_2S	H_2S, other inorganic compounds	H_2S, other reduced inorganic compounds, organic compounds
Photosynthetic oxygen	Yes	No	No
Principal source of carbon	CO_2	CO_2	CO_2 or organic compounds
Relations to oxygen	Aerobic	Strictly anaerobic	Strictly anaerobic or facultative aerobic

by a pinkish-violet or rose red scum. As much as 83 mg/L of pigment consisted mainly of bacterial chlorophylls and xanthophylls has been found in such pond water (24, 25). The major physiological distinctions between green plants and purple sulfur bacteria are shown in Table 8.1 (26).

1.2.2. Algal Population

Algal population varies from one pond to another. Algal counts can be as high as 15 million/mL. Green algae usually dominate because they can adapt better to environmental changes such as extreme temperature and dissolved oxygen. Raschke (27) has done a thorough study of algal population in a stabilization pond and found various species which are shown in Table 8.2.

In most ponds, coccoid green algae and green flagellates dominate the plankton throughout the year, while pennate diatoms and filamentous blue-green algae dominate the benthic flora. Green flagellates dominate in the winter and early spring, while coccoid greens, especially *Ankistrodesmus convolutus, Chlorella ellipsoidea,* and *Chlorella vulgaris* dominate in the fall, late spring, and summer. The population of some species such as *Chlorella*, *Chalamydomonas*, and *Euglena* may be reduced suddenly. Some evidence of zooplankton grazing and accumulation of inhibitory substance such as chlorellin provide partial explanation of the sudden (overnight) decrease of algal population in stabilization ponds. This algal autoinhibitor is possibly the same toxic substance that increases the rate of bacterial dieoff in lagoons.

1.2.3. Zooplankton and Insects

Protozoa and rotifers can be found abundantly following significant increase of bacterial and algal population. Their function in stabilization ponds is believed to be the same as in activated sludge and trickling filter processes in that they help to stabilize the prey population and therefore the treatment performance.

Usinger (28) found 52 species of aquatic insects in stabilization lagoons in California, whereas Kimerle (29) found 60 species in Central Missouri lagoons. The species compositions are similar in these two reports.

Midge larvae actively pump water in ponds; in doing so they remove settled algae and help in the circulation of large quantities of oxygenated water at the mud-water interface. Consequently they facilitate BOD removal in the pond. One or more of the three species of midges, *Glyptotendipes barbipes, Chironomus plumosus,* and *Tanypus punctipennis* comprise

Table 8.2
Algae found in tertiary wastewater stabilization pond and ditch influent

Division and order	Species
Chlorophyta	*Chlaniydomonas* supp.
Volvocales	*Chlamydomonas celerrirna* Pasch.
(Green algae)	*Chiarnydomonas tremulans* Skuja
	Chiorogonjum acus Nayal
	Chiorogonjum fusiforme Matw.
	Eudurina sp.
Ulotrichales	*Stigeolonium* sp.
Chlorococcales	*Ankistrodemus convolutes* Corda
(Green flagellated algae)	*Chlorella ellipsoidea* Gerneck
	Chlorella vulgaris Beyernick
	Coelastrum sp.
	Cruciogenia irregularis Willie
	Kirchneriella sp.
	Micractinium pusillum Fresenius
	Oocystis sp.
	Scenedesmus sp.
Euglenophyta	*Euglena* spp.
Euglenales	*Euglena pisciformis* Klebs
Chrysophyta	*Gomphonema parvulum* Küts.
Pennales	*Hontzschia* sp.
(Brown or yellow-Green diatoms)	*Navicula accomoda* Hust.
	Navicula cuspidate var. ambigua (Ehr.) Cleve
	Navicula excelsa Krasske
	Navicula gregaria Donk.
	Navicula kriegeri Krasske
	Navicula lanceolata (Ag.) Küts.
	Nitzschia spp. *Nitzschia accomodata* Hust.
	Nitzschia amphibia Grun.
	Nitzschia communia Rabh.
	Nitzschia diserta Hust.
	Nitzschia fonticola Grun.
	Nitzschia lateens Hust.
	Nitzschia palea (Kütz.) W. Smith
	Nitzschia thermalis Kütz.
Cyanophyta	*Oscillatoria amoena* (Kütz.) Gomont
Oscillatoriales	*Oscillatoria okeni*; Ag. Ex Gomont
(Blue-green algae)	*Oscillatoria tenuis* var. Natans Gomont
	Oscillatoria terebriformis Ag.

Table 8.3
Typical insect population in stabilization ponds

Type of lagoon	Organic Loading,[a] lb BOD/d/acre	Age yr	G. *barbipes* No./ft^2	*C. plumosus* No./ft^2	*T. punctipennis* No./ft^2
I	18.3	1.4	7913	37	9
II	32.0	3.0	1121	243	82
III	50.0	6.0	212	966	501

more than 94% of the total number of insects collected in the Midwestern lagoons (29). The number and the predominant species seem to vary with the type of lagoon, depending on age and organic loading. Table 8.3 shows such correlation.

Mosquito breeding takes place in any stabilization pond that provides a protected area for oviposition. Emergent vegetation, overflow structures, and swamp areas created by lagoon effluent are principal factors conducive to mosquito breeding. Predominant species found in lagoons by Kimerle (29) and Beadle (30) are *Culex pipiens* and *Cules tarsalis.*

Some benthic species of insects, such as *Planthemis lydia* and Laccophilus spp. serve as predators of midge larvae and pupae in stabilization ponds. With vegetation in poorly managed lagoons, predaceous insects may include Dytiscid larvae and adults, *Anax junius, Tropisternus lateralis nimbatus* larvae, *Belostoma* sp., *Odonata* naiads and *Hemiptera*. Relative densities and successive seasonal changes in aquatic insect populations are believed to be the result of climatic conditions as well as intraspecific and interspecific competition.

1.3. Classification of Stabilization Ponds

According to the type of biological transformation and methods of oxygen supply, stabilization ponds can be divided into four general classes (31).

1.3.1. Anaerobic Ponds

These are deep ponds where anoxic condition prevails throughout. Organic loadings are very high and BOD removal is limited to 80% or below. Further treatment of the anaerobic pond effluent by aerobic ponds is usually required.

1.3.2. Facultative Ponds

Facultative ponds receive medium to low organic loadings. Generally they are 8 ft deep or shallower. The bottom layer is usually anaerobic, but the surface layer is kept aerobic through photosynthesis and surface reaeration. BOD removal is higher than that of anaerobic ponds.

1.3.3. Aerated Lagoons

Oxygen supply in aerated lagoons for aerobic stabilization relies almost completely on mechanical aeration devices. Either air diffusion or mechanical aeration can be used. Depending on the power level used for aeration, aerated lagoons can be further classified as aerobic and aerobic-anaerobic lagoons (32). With power levels at 0.03 hp/1000 gal (6W/m^3) or above, ponds are usually aerobic throughout. At lower power levels, an anoxic bottom layer can be

expected. An aerobic-anaerobic lagoon is therefore equivalent to a facultative pond, except that the former uses mechanical aeration for oxygen supply. Aerated lagoons usually are deeper than facultative ponds and receive medium to high organic loadings. Treatment efficiency can be very high. The mixing characteristic that separates aerobic (near complete mix) to aerobic-anaerobic (poorly mixed) is more dependent upon the pond geometric parameters than upon operating parameters such as horsepower input. Detail discussion will be presented in Section 2, System Variables and Control.

1.3.4. High-Rate Aerobic Ponds

These are shallow ponds of only 12 to 18 in. working depth. Light penetration is essential to maximize algal production in which the bacterial-algal symbiosis affects BOD and nutrient removal. Very high organic loading is allowable with good BOD removal.

There are also stabilization ponds called tertiary ponds, multiple stage ponds, and integrated ponds. Considering the biological actions and methods of oxygen supply, these ponds or their components can fit into either one or different combinations of the four classifications previously described.

2. SYSTEM VARIABLES AND CONTROL

Traditionally, stabilization ponds are used for wastewater treatment because of their simplicity in construction and operation. Efforts at process control are very limited. Nevertheless, system variables of stabilization ponds can be identified through operational experience in the past several decades. This information is valuable to establish the criteria for the selection and design of different types of stabilization ponds, as well as to determine the control requirement for successful treatment.

2.1. Kinetics of Substrate Removal

The kinetics of substrate removal in stabilization ponds is less well defined compared to the activated sludge process, although bacterial oxidation is still primarily responsible for substrate removal. Except in the case of tertiary ponds, raw wastewater, including soluble and nonsoluble substrate, is introduced into the ponds. Some readily settle and exert oxygen demand later following an active anaerobic decomposition process of the sludge. The time lag depends on the degree of mixing in the pond and seasonal variation of temperature. Also unlike activated sludge processes where biological solid concentration is high and sludge age is under control, stabilization ponds are flow-through systems with low concentrations of biological solids. The sludge age is identical to the hydraulic detention time because no sludge recycle is practiced.

A general approach in considering substrate removal in a stabilization pond is to assume a first-order reaction for observed BOD_5 or ultimate BOD removal that applies to either overall solids or soluble solids only (33). A material balance around a pond taken as a complete-mix reactor will yield the following:

$$\frac{dS}{dt}(V) = QS_0 - QS - kSV \tag{13}$$

The equation states that the rate change of substance concentration S in a pond of liquid volume V is equal to the rate of substrate inflow (Q = flow rate, S_o = influent substrate concentration) minus the rate of substrate outflow (S = outflow substrate concentration, which is identical to the substrate concentration in the pond), minus the rate of substrate removal (k = substrate removal rate constant). At steady state, i.e., $dS/dt = 0$, Equation 13 becomes

$$\frac{S}{S_o} = \frac{1}{1 + k(V/Q)} = \frac{1}{1 + kt} \tag{14}$$

where t = hydraulic detention time.

Applying Equation 14 to multiple-stage ponds in series results in

$$\frac{S_i}{S_{i-1}} = \frac{1}{1 + k_i t_i} \tag{15}$$

$$\frac{S_n}{S_o} = \frac{1}{\sum_{i=1}^{n} (1 + k_i t_i)} \tag{16}$$

where S_i, k_i, and t_i represent respective parameters for pond i. Assuming all ponds in the series are of equal size, i.e., $v_i = V/n$, and that k_i is constant for all ponds in the series, it follows that;

$$\frac{S_n}{S_o} = \frac{1}{(1 + k_i t_i)^n} \tag{17}$$

It can be easily seen that operating stabilization ponds in series is advantageous because the total volume required for a series of ponds is considerably less than that required for a single pond. Table 8.4 after Metcalf and Eddy Inc. (34) can be used to illustrate this point.

Table 8.4
Required reactor volume for complete-mix ponds

	Required reactor volume,[a] Q/k			
No. of reactors in series	85% Removal efficiency	90% Removal efficiency	95% Removal efficiency	98% Removal efficiency
1	5.67	9.00	19.00	49.00
2	3.18	4.32	6.96	12.14
4	2.48	3.10	4.48	6.64
6	2.22	2.82	3.90	5.50
8	2.16	2.64	3.60	5.04
10	2.10	2.60	3.50	4.80
Plug flow	1.90	2.30	3.00	3.91

[a]Volume of individual reactors equals value in table divided by the number of reactors in series.

The table also indicates that the volume differential becomes more pronounced as the removal efficiency increases. However, one should view with reservation the beneficial use of multiple-stage ponds against one single pond. The fallacy lies in the assumption of a constant specific substrate removal rate k for all ponds in the series. The substrate biodegradability changes as the wastewater moves down the ponds. Settleable substrates are readily removed in the first few ponds and seldom if ever enter the last few ponds in the series. The k rate should decrease and consequently should greatly offset the volume differential that Table 8.4 tends to indicate. This is analogous to the decreasing k value in trickling filter operation with the effluent recirculated many times passing through the filter bed. There is evidence that fractionation of the total volume of a pond to smaller ones in a series according to the concept of complete mixing and a first-order reaction does not result in any advantage. Fleckseder (32) reported in an aerated lagoon study that the staging of the pond did not improve the quality of effluent from the system. The effluent quality, both settled and filtered, from a three-pond system, after equal detention time, was the same, within the precision of the measurements, as those from the two-pond and one-pond systems.

It is important to note that Equation 14 was developed based on complete-mix flow condition. A stabilization pond is neither a complete mix nor a plug flow system, but rather an intermediate system. Wehner and Wilhem (35) have developed an equation for a chemical reactor with a value of diffusivity constant between zero and infinity (true plug flow and complete-mix conditions respectively). The Wehner-Wilhem equation was developed based on first-order reaction and therefore could apply to stabilization ponds (8):

$$\frac{S}{S_o} = \frac{4ae^{(1/2d)}}{(1+a)^2 e^{(a/2d)} - (1-a)^2 e^{-(a/2d)}} \tag{18}$$

$$d = \frac{D}{UL} = \frac{Dt}{L^2} \tag{19}$$

where, $a = \sqrt{1+4kd}$, d = diffusivity constant or dispersion number (dimensionless), D axial dispersion coefficient (ft^2/h), U = fluid velocity (ft/h) and L = characteristic length or travel path of a typical particle in the reactor (ft).

Thirumurthi (36) has prepared a chart for the Wehner and Wilhem Equation that is plotted here as Figure 8.4. Either Equation 18 or Figure 8.4 can be used for design purposes. There is a common problem associated with the use of Equations 14 and 18 for the design of stabilization ponds. It is very difficult to make an intelligent selection of the values of k, which are affected by temperature, influent waste quality, nutrient availability, organic loading, and other biological factors. Thirumurthi (37) has provided some evidence, based on the analysis of the operation data of several unaerated, facultative ponds, that a standard BOD removal coefficient k of $0.056\,d^{-1}$ can be assigned to the ponds if the following standard environment prevails: (a) pond temperature at 20°C, (b) organic load of 60 lb/d/ac. (677 kg/d/ha), (c) absence of industrial toxic elements and benthic load, (d) minimum visible solar energy at the rate of 100 Langleys/d. A correction factor C_o for different organic

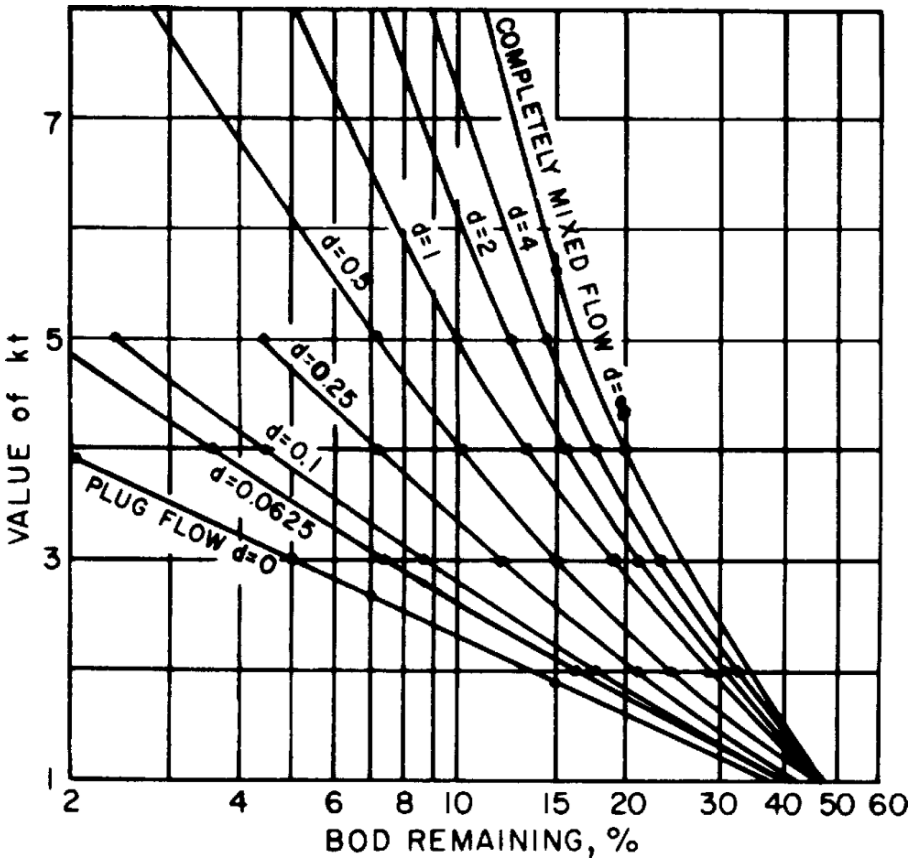

Fig. 8.4. Design chart for chemical reactor using Wehner and Wilhem equation after Thirumurthi.

loading also has been suggested as:

$$C_o = 1 - \frac{0.083}{k}\left[\log\frac{60}{L}\right] \tag{20}$$

where L = organic load in lb/ac./d to be used in the design. Field BOD removal coefficient k may vary from 0.01 to 1.28/d. The selection of proper values of d poses an additional problem to the use of Equation (18). Although a range of 0.1 to 2.0 is suitable for most stabilization ponds, a more precise determination of a proper d value could only be made by a tracer study in an existing pond. The value of d represents short-circuiting in ponds, exit and entrance hydraulic devices, geometry of ponds, and other hydraulic mixing characteristics. Levenspiel et al. (38) have outlined a procedure to determine the value of d in a tracer study. By adding a known amount of tracer into the pond and with a constant flow through the pond, the exit tracer concentration over a long period of time (20 to 30 d) is measured. A curve of tracer

concentration versus time can be plotted. The following equations are then used to determine the values of d and actual detention time t that appears in Equation 18

Mean (actual) detention time

$$t = \frac{\sum tC}{\sum C} \tag{21}$$

$$\sigma_t^2 = \frac{\sum t^2 C}{\sum C} - \left(\frac{\sum tC}{\sum C}\right)^2 \tag{22}$$

$$\sigma^2 = \sigma_t^2/t^2 = 2d - 2d^2(1 - e^{-1/d}) \tag{23}$$

The term, d, can be calculated by trial and error from Equation 23 in which σ_t = standard deviation and σ = Variance.

2.2. Oxygen Supply

Except in anaerobic ponds and the bottom of facultative ponds, oxygen is needed for bacterial oxidation of waste organics. Major sources of oxygen supply in practice are (a) mechanical aeration and (b) photosynthetic oxygenation. Surface reaeration is normally insignificant because of smaller size of ponds compared to lakes and streams.

The theoretical oxygen demand can be estimated from the BOD to be removed and the daily production of volatile suspended solids in the pond. Readers should refer to the Section, Aeration Requirements and Process Design, Example 3 in Activated Sludge Processes, Chapter 6 in this book, for a pertinent discussion. An example of process design including the aeration requirement is given later in this chapter.

Oxygen is also consumed in the presence of H_2S gas emitted from zones of anaerobic decomposition in the pond.

$$H_2S + 2O_2 \xrightarrow{\text{bacteria}} H_2SO_4 \tag{24}$$

The oxygen demand to satisfy sulfide oxidation could be a significant portion of the total oxygen demand in the pond. It depends on the sulfide concentration in the pond water and the degree of odor control to be considered in design or in operation. A design example pertinent to this subject is given later in this chapter.

The benthic oxygen demand does not play a significant role in design estimation of oxygen supply. The problem naturally does not exist in high-rate aerobic ponds and/or aerobic ponds with good mixing keeping the sludge in suspension most of the time. There is no intention to meet all the oxygen demand in other types of stabilization ponds and therefore benthic oxygen demand can be neglected.

In high-rate aerobic ponds and in the aerobic zone of a facultative pond, the oxygen production through photosynthesis is equated to the oxygen demand. The solar radiation varies with the time of year and the latitude. Table 8.5 gives the typical values of visible solar energy for each month and for various latitudes (39).

Not all the available solar energy can be fixed by algal cells. The overall photosynthetic efficiency usually lies in between 0.5% to 6.0%. Several environmental factors play an impor-

Table 8.5
Probable values of visible solar energy as a function of latitude and month

Latitude		Month[b]											
		Jan.	Feb.	Mar.	Apr.	May	Jun.	Jul.	Aug.	Sept.	Oct.	Nov.	Dec.
0	max	255	266	271	266	249	236	238	252	269	265	256	253
	min	210	219	206	188	182	103	137	167	207	203	202	195
10	max	223	244	264	271	270	262	265	266	266	248	228	225
	min	179	184	193	183	192	129	158	176	196	181	176	162
20	max	183	213	246	271	284	284	282	272	252	224	190	182
	min	134	140	168	170	194	148	172	177	176	150	138	120
30	max	136	176	218	261	290	296	289	271	231	192	148	126
	min	76	96	134	151	184	163	178	166	147	113	90	70
40	max	80	130	181	181	286	298	288	258	203	152	95	66
	min	30	53	95	125	162	173	172	147	112	72	42	24
50	max	28	70	141	210	271	297	280	236	166	100	40	26
	min	10	19	58	97	144	176	155	125	73	40	15	7
60	max	7	32	107	176	249	294	268	205	126	43	10	5
	min	2	4	33	79	132	174	144	100	38	26	3	1

[a] After Table II, W. J. Oswald and H. B. Gotaas, "Photosynthesis in Sewage Treatment" *ASCE Proceedings* 81, Separate No. 686 (May 1955).

[b] Values of S in Langleys, cal/(cm^2)(d)

Correction for cloudiness:

$$S_c = S_{min} + r(S_{max} - S_{min})$$

where r = total hours sunshine/total possible hours sunshine

Correction for elevation up to 10,000 ft:

$$S_c = S(1 + 0.01e)$$

where e = elevation in 10 feet.

tant role in determining the photosynthetic efficiency. Oswald (40) presented the appropriate correction factors for the determination of the photosynthetic efficiency E, as shown in Table 8.6.

In Table 8.6, $S = 1000\,S/d$ where S is the visible solar radiation energy from Table 8.5 and d is depth of the pond, cm. The overall photosynthetic efficiency E then can be calculated as;

$$E = \frac{F_s + F_d + F_p + F_L}{4} \cdot T_c \tag{25}$$

with the critical photosynthetic efficiency necessary to provide enough O_2 to meet the BOD loading given by Oswald (40) as:

$$F_c = 0.94\frac{Ld}{St} \tag{26}$$

Table 8.6
Percent photosynthetic efficiencies as a function of environmental factors

Light energy flux S		Detention Period D		Diurnal Illumination P		Applied BOD L 5d, 20°C		Temperature T	
Cal/d/L $\times 10^{-2}$	F_s	d	F_d	%	F_p	mg/L	F_L	°C	T_c
7.2	7.4	1.0	6.2	30	–	10	0.2	4	0.01
14.4	6.9	1.5	5.7	33	–	25	0.5	6	0.02
21.6	6.7	2.0	4.9	36	–	50	1.2	8	0.23
28.8	6.4	2.5	4.3	39	4.8	75	2.1	10	0.49
36.0	6.2	3.0	3.8	42	4.5	100	2.6	12	0.70
43.0	6.0	3.5	3.5	45	4.1	125	3.1	14	0.82
58.0	5.7	4.0	3.1	48	3.9	150	3.6	16	0.91
72.0	5.3	4.5	2.9	51	3.7	175	4.0	18	0.96
108.0	4.6	5.0	2.7	54	3.5	200	4.3	20	1.00
144.0	4.0	5.5	2.5	57	3.4	250	4.8	22	0.99
180.0	3.5	6.0	2.3	60	3.3	300	5.1	24	0.96
206.0	3.0	7.0	2.1	66	3.1	350	5.3	26	0.92
290.0	2.2	8.0	1.9	72	3.0	400	5.4	28	0.87
360.0	1.6	10.0	1.6	78	3.0	500	5.6	32	0.73

The rate at which solar energy is used /cm^2 of pond area in algal production can then be expressed as:

$$hW_a = ESA \tag{27}$$

in which W_a = net weight of algal cells synthesized daily, g/d; S = solar radiation in Langleys, cal/cm^2/d; A = surface area of pond in cm^2; h = unit heat of combustion, cal/g, and expressed as a function of degree of reduction of cellular organic material (41) or,

$$h = 400 + 127R$$
$$h = 400 + 127\frac{100[2.66(\%C) + 7.94(\%H) - (\%O)]}{398.9} \tag{28}$$

The average algal cell composition can be taken as $C_7H_{8.1}O_{2.5}N$ in stabilization ponds. The amount of oxygen yield, W_{O_2}, in g/d, can now be estimated as:

$$W_{O_2} = pW_a \tag{29}$$

in which p = oxygenation factor, or the amount of oxygen produced per unit weight of algal cell synthesized. $C_7H_{8.1}O_{2.5}N$, p is equal to 1.67.

A much simpler method to estimate photosynthetic oxygenation suggested by McGauhey (42) is to use the following equation;

$$W_{O_2} = 0.25FS \tag{30}$$

in which F = oxygenation factor, S = solar radiation in cal/cm^2/d and the oxygen yield W_{O_2}, is expressed in lb O_2/ac./d. It should be noted that the oxygenation factor, F, here has a different meaning in that it is a proportional coefficient relating the oxygen yield to solar radiation. Comparing Equation 30 with Equations (25) through (29), it can be seen that F is a function of many environmental factors that determine the overall photosynthetic efficiency and the unit heat of combustion h. There is no rational way to determine the value for F in Equation (30) at the present time, although McGauhey (42) recommended a value of 1.6 for approximately 90% BOD_L removal. One should therefore exercise his judgment carefully in using Equation 30 for stabilization pond design.

After the photosynthetic oxygenation is evaluated, one can proceed to equate W_{O_2} to BOD to be removed by the pond. A design example will be given later in this chapter.

2.3. *Temperature Effect*

2.3.1. Temperature Coefficient

As in all biological treatment systems, the effect of temperature on the reaction rate in stabilization ponds is generally expressed by the modified Arrhenius equation;

$$k_T = k_{20}\theta^{(T-20)} \tag{31}$$

The value of the temperature coefficient varies with the type of system. A range of 1.0 to 1.03 has been reported for activated sludge process and 1.02 to 1.04 for trickling filter process. A higher value is generally accepted for stabilization ponds, e.g., 1.06 o 1.07 for aerated ponds.

Two important considerations are lacking in choosing the θ value for stabilization pond design. Some ponds receive raw wastewater with significant amounts of settleable organic solids in addition to soluble organics. The θ values for activated sludge and trickling filter processes apply mainly to soluble organic removal. Studies conducted in Texas (38, 43, 44) provide evidence that θ for the oxidation of soluble BOD can be taken as 1.036 to 1.04. However, anaerobic zones are established in many ponds and it is known that anaerobic decomposition is much more sensitive to temperature change than aerobic oxidation, it is therefore suggested that different θ values should be used according to Table 8.7.

2.3.2. Heat Loss and Heat Conservation

Heat loss is relatively high in stabilization ponds compared to other biological treatment processes because of the large surface area in relation to pond volume. Heat loss consists of

Table 8.7
Temperature coefficient for stabilization ponds

Type of stabilization	Suggested θ values
Aerobic lagoon, high-rate aerobic ponds, tertiary ponds	1.03–1.04
Facultative ponds, aerobic-anaerobic ponds	1.05–1.06
Anaerobic ponds	1.06–1.085

evaporation H_e, convection H_c, and radiation H_r, according to Barnhart (44)

$$H = H_e + H_c + H_r \tag{32}$$

where

$$H_e = 0.00722\ H_v C(1 - 0.1W)(V_w - V_a) \tag{33}$$

$$H_c = (0.8 + 0.32\ W/2)(T_w - T_a) \tag{34}$$

$$H_r = 0.1(T_w - T_a) \tag{35}$$

in which H_v is the latent heat of vaporization; C is a constant characteristic of the pond; W is mean wind velocity; V_w is vapor pressure at water surface, in. Hg; V_a is vapor pressure in atmosphere, in. Hg; T_w is pond water temperature, °F; T_a is air temperature, °F. The net heat gain by solar radiation cannot be defined from available data although Table 8.5 provides some quantitative estimation. A material balance around a pond taken as a complete-mix reactor can yield the information on pond water temperature once the influent flow and influent temperature T_i are specified. A simplified equation has been presented by Mancini et al. (43) as follows:

$$\frac{Q}{A} = \frac{f(T_w - T_a)}{(T_i - T_w)} \tag{36}$$

where Q is the flow rate in MGD, A is surface area in ft^2. All temperature terms are expressed in °F, f is a proportionality factor that incorporates all appropriate heat transfer coefficients and includes the effects of wind, humidity, and surface area increase caused by aeration if applicable. A typical f value for the Eastern United States is given as $1.4 - 12 \times 10^{-6}$ and for the Midwestern United States, $21.6 - 85.5 \times 10^{-6}$. Because stabilization pond performance is more sensitive to temperature change, it is a good practice to conserve heat energy in the pond, particularly when facultative ponds and anaerobic ponds are used. Equation (36) can be used in the design of stabilization ponds to minimize heat loss. An example is included later in this chapter.

2.4. Detention Time

Operated as a flow-through system without sludge recycle the effluent organic concentration is a function of the hydraulic detention time. Equations (14) and (15) both show the relationship between detention time and other system variables in determining the effluent organic concentration. By examining Equation (14) or Figure 8.4, it is clear that with a given system, the product kt is fixed when the allowable BOD remaining is specified. An unnecessarily prolonged detention time (oversized pond), resulting in decreased organic load and reduction of k rate, has no net gain in treatment efficiency.

Hydraulic detention has an indirect affect in the treatment efficiency of a stabilization pond in that the overall photosynthesis efficiency, E, decreases gradually with increasing t as shown in Table 8.6. The effect is, however, relatively insignificant compared to the effect of temperature and applied BOD load on photosynthesis efficiency.

3. DESIGN CRITERIA

3.1. Design Parameters

Because of the highly complex nature of the ecological system in stabilization ponds and the minimal efforts in operational control, the method of design is not as well defined as in other biological treatment processes. Based on the operating experience of pond treatment systems, the design criteria are listed in Table 8.8. The design criteria for any given geographical location reflect the influence of prevailing climatological conditions. In general, organic loadings are higher and detention time is shorter for milder climate areas. Wastewater characteristics and effluent quality also are important considerations. Consequently ranges of values are given for each listed design parameter.

A survey of facultative pond design criteria in the 50 states by Canter (45) provides results summarized in Table 8.9. A very wide range of detention times is reported for the apparent reason that facultative ponds with ice cover usually cease to function and therefore the total flow in winter should be accommodated without outflow from the pond.

3.2. Inlet Structures

A single pipe inlet is preferred for small ponds where no mixing occurs. Such an inlet should extend to the center of the pond so that the sludge can be distributed by wind induced currents. This requires subsurface location of the pipe, usually along the bottom of the pond with the top of the pipe being just below the average elevation of the bottom. Gravity feed inlet at a 2% slope should extend past the toe of the berm. By providing a saucer-shaped excavation in front of the pipe, a blocking of the pipe by sludge can be avoided. The depression should extend not more than one pipe diameter plus 1.0 ft below the average bottom elevation according to Dawson (46). Force main inlet should terminate with an upturned elbow. A concrete spill pad of no smaller than 2.0 ft^2 is recommended at the end of the discharge line. Manholes are recommended where pipes pass through the embankment to facilitate inspection and maintenance.

For larger ponds without artificial mixing, a single inlet pipe presents the problem of possible overloading in the feed zone, allowing odor to develop. A multiple-entry scheme should thus be devised. Discharge velocities at 8 ft/s or higher from a multiple-entry system can provide sufficient mass pond movement to induce internal pond circulation and mixing. The pipes of the multiple-entry scheme need not be extended well into the pond. If lining is not provided on the inside embankment, a rip rap is recommended at the end of the discharge pipes.

3.3. Outlet Structures

Both overflow and drain outlets should be provided. A pond drain that can empty the pond within 2 days should be included. A mild slope in the pond bottom in all directions toward the drainage point should be provided.

An overflow weir structure can be located on the windward side, with removable planks to adjust the water level. The weir could be placed inside a standard manhole in the embankment.

Table 8.8
Design criteria for stabilization ponds

Criteria	Anaerobic pond	Facultative pond	Aerated ponds		
			Aerobic-anaerobic	Aerobic	High-rate aerobic
Organic loading, lb BOD_5/ac./d	200–1000	15–75[a]	30–100	800–2500	60–200
Detention time, d	20–180	20–180[b]	10–20	3–10	3–6
Depth, ft	8–15	3–6[b]	3–8	6–20	2–4
BOD_5 conversion, %	50–85	80–95	80–95	80–95	80–95
Application	Treatment of untreated wastewater and sludge	Treatment of untreated to partially treated wastewater	Treatment of untreated to partially treated wastewater	Treatment of primary effluent; tertiary pond	Treatment of soluble organic; nutrient removal; tertiary pond
Effluent characteristic					
BOD concentration, mg/L	100–300[a]	20–50	20–50	20–30	20–30
Suspended solids, mg/L	80–160	160–400	100–350	250–300	150–350
Algae concentration, mg/L	—	40–160	10–40	—	80–200

[a] Usually followed by a facultative or aerobic pond for further treatment.
[b] In arctic and sub-arctic regions, 20 lb/ac./d organic loading, 6–8 months detention time, and 6 to 8 ft deep are recommended by Dawson (46).

Table 8.9
Design criteria for facultative ponds in United States

	Region		
Variable	North[a]	Central	South
Number of states	18	17	15
Organic loading, lb BOD_5/ac/d			
Mean	26	33	44
Range	16.7–40 [5]	17.4–80 [1]	30–50 [2]
Median	21	33	50
Loading, population/ac.			
Mean	124	189	267
Range	100–200 [7]	100–400 [4]	175–300 [3]
Median	100	200	295
Detention time, d			
Mean	117	82	31
Range	30–180 [1]	25–180 [5]	20–45 [9]
Median	125	65	31

[a] Numbers in [0] indicate the number of states for which no value was obtained. North region, north of 42° latitude; center region, between 37° to 42° latitude; south region, south of 37° latitude.

The overflow line should be at least 6 in. in diameter and must be used with a gate-valve equipped with a valve box and a handle extended upstream of the weir.

3.4. Transfer Pipes

Transfer pipes or interconnecting piping for multiple pond systems should be sized such that the head loss with pipes flowing about 2/3 to 3/4 full does not exceed 3 to 4 in. Supply-and-return-channel sizing should be such that total channel loss is no more than one-tenth of the transfer-pipe losses. In this way, uniform distribution to all pond areas at all recirculation rates, as well as water-surface continuity between all points in the system, can be maintained (47).

3.5. Berm Design

Berms or dikes should be constructed of impervious material and compacted to form a stable structure. A compacted clay core for water proofing is recommended. Sealing with an asphaltic or polyethylene membrane is an alternative to a clay core. The top width of the berm should be a minimum of 10 ft wide if vehicle access is desirable. Most commonly used side slopes are three horizontal to one vertical. Steeper slopes can be used if the structural properties of the soil permit. A shallower slope at 6:1 is recommended when wind erosion is a factor. A minimum 2-ft summer freeboard is advisable; this can be reduced to 1 ft during the winter (when ice cover is expected) to increase the storage volume. If a lining is not used, the berm should be protected from at least 1 ft below the minimum water surface to at least 1 ft above the maximum water surface. Use of a perennial type, low-growing, spreading grass for berm protection above the water level has been widely practiced.

3.6. Bottom Preparation

All vegetation and organic topsoil should be removed. Organic material removed during this process should not be used for core material in dike construction. To avoid leaking that causes groundwater contamination and short-circuiting to a receiving stream, the bottom should be made impermeable. Although sealing with bentonite or asphalt has been practiced, use of a lining is a more acceptable practice. An impermeable lining of PVC or Hypalon synthetic rubber can be installed for the entire pond (i.e., bottom and sides of embankment).

For most stabilization ponds, a 20 to 30 mil thick liner, nonreinforced type can be used. The liner is usually prefabricated with leakproof seals and installed on site over a smooth bottom surface free of any obstacles that could puncture the liner. In some states, perforated pipe monitoring systems have to be laid under the lining. The perforated pipe also serves as vent and drainage system for gases and water built up under the liner. Berm trenching is required for anchoring the liner, and usually a 1-ft earth cover is required to protect the liner.

4. PRACTICE AND PROBLEMS IN PROCESS CONTROL

Although stabilization ponds are simple devices for wastewater treatment, the basic objectives should be upheld for providing an effective treatment according to design specifications and not creating objectionable conditions or public health hazards in the vicinity or downstream of the receiving waters. Common practices in process control and some problems associated with them are discussed in this section. Remedies for the control problems are also included.

4.1. Staging of Ponds

A common practice in stabilization ponds with mixing is to fractionate the total design volume to smaller ponds. The validity of belief in increasing the treatment efficiency by the staging of ponds has been examined previously in this chapter. Staging of ponds also is not an economical mode of land use because a single pond occupies less land area per unit volume. In addition, the staging of ponds adds to the construction costs in excavation, interconnecting pipes or channels, and requires more maintenance.

Despite the disadvantage of pond fractionation, it adds to the operational flexibility by recirculation, minimizes upset and nuisance, and maximizes stability. This is a tradeoff between a less expensive, simple-in-operation unit and a more expensive, reliable, sophisticated pond system. The decision usually is in favor of staging the pond if the system is designed for (a) efficient treatment of raw or primary wastewater to produce an effluent quality compatible with other types of secondary treatment process and (b) flexibility in being able to switch operations from series to parallel or their combinations so that various organic and hydraulic loadings can be accommodated as seasonal load changes.

4.2. Pond Recirculation

Recirculation can be practiced for a single pond or multiple-pond system. It dilutes the wastewater influent and incorporates some photosynthetic oxygen to the feed zone so that

odors and anaerobic conditions can be avoided. For n number of ponds in a series, the dilution can be expressed as:

$$S_d = \frac{S_o}{1+r} + \left(\frac{r}{1+r}\right) S \tag{37}$$

in which S_d is the substrate concentration in the diluted wastewater entering the first pond, S_o is the influent substrate concentration, r is the recycle ratio or the ratio of recycled flow rate R to influent flow rate Q, and S is the effluent substrate concentration from the nth pond. As r increases, S_d becomes smaller and the incoming organic load is spread more evenly throughout the pond system. The overloading of the first eliminated completely. Because the design incorporates low head losses in interconnecting pipes and channels, recirculation can be accomplished with high volume, low-head propeller pumps. Caldwell et al. (47) recommend siphon discharge flap gates. An auxiliary pump with an air eductor maintains the siphon. Siphon breaks are provided to insure positive backflow protection. Multiple- and/or variable-speed pumps can be used to adjust the recirculation rate to seasonal load changes.

4.3. Pond Mixing and Aeration

Mixing and aeration are accomplished concurrently when compressed air diffusers or mechanical aerators are used. In high-rate aerobic ponds in which intermittent mixing is recommended to eliminate a sludge buildup at the bottom, only a high-volume, low-head propeller pump is needed because aeration is not necessary in the pond operation.

Where aeration is of prime consideration, diffused aeration is used in ponds deeper than 10 ft. The advantage of diffused aeration is that air can be distributed evenly over the entire pond of any geometry. It is preferable to mechanical aeration in northern areas, because ice buildup in the latter might be excessive. Proprietary devices have been developed to introduce compressed air through slits in piping laid on the pond bottom. Bubble-gun aerators also combine good mixing characteristics with aeration. Diffused aeration is also advantageous in that tapered aeration can be incorporated easily so that air supply can be more efficiently used to meet the oxygen demand at different locations of the pond.

Mechanical aerators are generally divided into two types: cage aerators and the more common turbine and vertical-shaft aerators. The former are suitable for shallow ponds of 5-ft depth or less, while the latter require a minimum depth for economical use that depends on the horsepower of the unit. The cage aerator appears to have a greater pumping capacity than the propeller aerator, because it does not recycle much the volume pumped. Mechanical aerators can be installed on platforms in a fixed position or the floating type can be mounted out in the pond, and spaced about to provide even distribution. Floating cage aerators may be mounted either in the pond or directly off the dike slopes. When mounted off the dike slopes, they can be close to the pond inlets. Better oxygen supply on the inlet side and easy access for maintenance and repair are the advantages of such arrangements.

Commercial-size surface aerators available range in efficiency from 2 to 4 lb O_2/hp/hr and turbine aerators from 2 to 3 lb O_2/hp/hr. Efficiency claims are usually based on standard testing conditions of 20°C, zero dissolved oxygen concentration in tap water and a nonsteady state condition. For design purposes, the standard oxygen transfer rate, N_o, must be adjusted

Table 8.10
Solubility of oxygen (mg/L) at various temperatures and elevations (Based on sea level barometric pressure of 760 mm Hg)

	Elevation, ft above sea level						
Temperature °C	0	1000	2000	3000	4000	5000	6000
0	14.6	14.1	13.6	13.2	12.7	12.3	11.8
2	13.8	13.3	12.9	12.4	12.0	11.6	11.2
4	13.1	12.7	12.2	11.9	11.4	11.0	10.6
6	12.4	12.0	11.6	11.2	10.8	10.4	10.1
8	11.8	11.4	11.0	10.6	10.3	9.9	9.6
10	11.3	10.9	10.5	10.2	9.8	9.5	9.2
12	10.8	10.4	10.1	9.7	9.4	9.1	8.8
14	10.3	9.9	9.6	9.3	9.0	8.7	8.3
16	9.9	9.7	9.2	8.9	8.6	8.3	8.0
18	9.5	9.2	8.7	8.6	8.3	8.0	7.7
20	9.1	8.8	8.5	8.2	7.9	7.7	7.4
22	8.7	8.4	8.1	7.8	7.7	7.3	7.1
24	8.4	8.1	7.8	7.6	7.3	7.1	6.8
26	8.1	7.8	7.6	7.3	7.0	6.8	6.6
28	7.8	7.5	7.3	7.0	6.8	6.6	6.3
30	7.5	7.2	7.0	6.8	6.5	6.3	6.1
32	7.3	7.1	6.8	6.6	6.4	6.1	5.9
34	7.1	6.9	6.6	6.4	6.2	6.0	5.8
36	6.8	6.6	6.3	6.1	5.9	5.7	5.5
38	6.6	6.4	6.2	5.9	5.7	5.6	5.4
40	6.4	6.2	6.0	5.8	5.6	5.4	5.2

to reflect anticipated field conditions by using the following relationship;

$$\text{Oxygen transfer rate at field condition, } N = N_0 \frac{(C_s\beta - C_r)(1.024)^{T-20}\alpha}{9.17} \tag{38}$$

in which both N and N_o are expressed in lb O_2/hp/h, β = oxygen saturation coefficient in wastewater, usually 0.95, C_s = oxygen saturation in tap water at a given temperature and altitude, as given in Table 8.10. Correction also can be made for the salinity of the water, as given in Table 8.11, T = temperature, and α = oxygen transfer rate correction factor for wastewater, between 0.8 and 0.9 depending on the strength of the wastewater, C_r = oxygen concentration to be maintained in the wastewater.

A mathematical model developed by Wang (48) can be used to calculate the dissolved oxygen concentration in water at given barometric pressure, water temperature and chloride concentration.

The oxygen requirement can be estimated from the amount of BOD to be removed or in general from a theoretical amount of 1.5 lb O_2/lb of BOD removed. The application of Equation (38) and the design of an aerator are illustrated later in this chapter in Examples of Process Design.

Table 8.11
Solubility of oxygen (mg/L) at various temperatures and salinity (Based on a Barometric Pressure of 760 mm Hg)

	Salinity, %					
Temperature °C	0	4.0	8.0	12.0	16.0	20.0
0	14.6	13.9	13.2	12.5	11.9	11.3
2	13.8	13.2	12.5	11.9	11.4	10.8
4	13.1	12.5	11.9	11.3	10.8	10.3
6	12.4	11.8	11.3	10.8	10.3	9.8
8	11.8	11.3	10.8	10.3	9.8	9.4
10	11.3	10.8	10.3	9.8	9.4	9.0
12	10.8	10.3	9.8	9.4	9.0	8.6
14	10.3	9.9	9.4	9.0	8.6	8.3
16	9.9	9.4	9.0	8.6	8.3	8.0
18	9.5	9.1	8.7	8.3	8.0	7.6
20	9.1	8.7	8.3	8.0	7.7	7.4
22	8.7	8.4	8.0	7.7	7.4	7.1
24	8.4	8.1	7.7	7.4	7.1	6.9
26	8.1	7.8	7.5	7.2	6.9	6.6
28	7.8	7.5	7.2	6.9	6.6	6.4
30	7.5	7.2	7.0	6.7	6.4	6.2
32	7.3	7.0	6.7	6.5	6.2	6.0
34	7.1	6.8	6.5	6.3	6.0	5.8
36	6.8	6.6	6.3	6.1	5.8	5.6
38	6.6	6.4	6.1	5.9	5.6	5.4
40	6.4	6.2	5.9	5.7	5.4	5.2

4.4. Odor Control

Earlier in this chapter, in the discussion of pond ecology and process reactions, it was mentioned that anaerobic zones in stabilization ponds can generate odorous compounds such as H_2S and mercaptans that escape the pond. The oxygenated upper layer normally serves as a protective shield in keeping these odorous endproducts of anaerobic decomposition from escaping the pond. This is because the compounds are chemically or biochemically oxidized and rendered innocuous before they reach the surface. During periods of upset, i.e., when high H_2S concentration exists, algal growth and the associated oxygen production may be inhibited. Odorous compounds may reach the atmosphere as a result.

Other sources of odor are the endproducts of the metabolism of certain blue-green algae, diatoms and pigmented flagellates, and also the oil products released by dead algal cells. These sources are highly unpredictable and generally are not considered a major odor problem.

Gloyna et al. (12) have developed two empirical equations for predicting the sulfide concentrations in facultative ponds based on several variables:

$$S^{2-} = a\ (\text{BOD surface load}) + b\ (SO_4^{2-}\ \text{influent conc.}) + c\ (\text{detention time}) + d \qquad (39)$$

$$S^{2-} = a'(\text{BOD surface load}) + b'(SO_4^{2-}\ \text{surface load}) + c'(\text{detention time}) + d' \qquad (40)$$

in which S^{2-} is sulfide in mg/L; BOD surface load in lb/ac./d, from 68 to 136; SO_4^{2-} influent concentration in mg/L, 23 to 200; SO_4^{2-} surface load in lb/ac./d, 11–188; and detention time in days, 15–30. The values of the coefficients a, b, e, d, a', b', c', and d' were developed in laboratory experiments. The results led to the following:

$$S^{2-} = (SO_4^{2-})[0.0001185(\text{BOD surface load}) - 0.001655(\text{detention time}) + 0.0553] \quad (41)$$

in which S^{2-} is 24-h average sulfide concentration in the pond and SO_4^{2-} is concentration of sulfate ion in the influent. The sulfide could be oxidized before escaping the pond if sufficient oxygen is provided. An incorporation of this equation in the process design will be presented later.

4.5. Algae Removal

The suspended solids concentrations in the stabilization ponds effluent given in Table 8.8 are much higher than those of activated sludge or trickling filter processes unless a physical solid-separation step is included in the pond system. Because the discharge of suspended solids contributes BOD and nutrients to receiving waters, a reasonable amount of solids removal from the effluent is desirable and, in fact, is required by many states' regulations for high-quality receiving waters. Table 8.8 shows that algal cells in the effluent constitute only a part of the suspended solids. Even in high-rate aerobic ponds, algal production is only 50% to 60% of the effluent's suspended solids. Any physical or chemical separation process will remove both algal cells and other microbial solids.

Many processes have been investigated for the removal of algal cells from suspended solutions. The important features and operations of the most promising ones are summarized in Table 8.12.

There are other investigations on algal cell removal using ion exchange columns (51, 59) or sand filtration. These processes are either not economical and/or lack confirming field studies of their results. It would be advisable to find a market for algae harvested from stabilization ponds. Although the most important benefit now is the reduction of the quantity of organic and inorganic nutrient by removal of algae from the pond effluent, other algae removal's application has been explored. The harvested algae can be used as a livestock feed because it has high nutrient and protein content, and it is reported to be highly acceptable to livestock (58). Another potential use of the cultivated algae is to produce biomass to be used for distilling alcohol fuels, or producing methane gas.

4.6. Insect Control

With improper control or lack of consideration in design, stabilization pond environments often become a nuisance and public health hazard. Of primary importance are the breeding places that ponds and their effluents can provide for mosquitoes. Many of the predominant species of mosquitoes found in ponds are the primary vectors of encephalitic diseases (30). Their breeding takes place in any pond that provides a protected area for oviposition. Egg rafts and larvae can be found along the shoreline in vegetation, in water-filled gullies created by wheels of mowing equipment at the water's edge, in some overflow structure, in floating sludge that is allowed to accumulate in corners (29. Overflow boxes also harbor mosquito egg rafts and larvae because when water is not flowing through such boxes they provide a

Table 8.12
Algal removal processes

Process	Field or laboratory evaluation	Economic evaluation	Reference
Froth flotation. Fine bubble aeration, foam collection, pH adjustment, neutralization	Laboratory and field evaluation	Limited, claimed economical use	47, 49, 50, 52, 53
Chemical precipitation: Alum, lime, or cationic polymers with pH control, sedimentation	Laboratory pilot, and field-scale studies	Not extensive, authors claimed economical use	47, 51, 53
Centrifugation: for both concentration and dewatering	Laboratory, pilot, and field studies	Marginal economical use	51
Mechanical separation: recoated with paper pulp, microstrainer, vacuum, heated, air-dried to form algal paper, shredded to form animal feed	Laboratory study	Not available	47
Submerged rock filter: upflow mode	Laboratory study with pond effluent	Limited. Economical for small facility	54
Upflow Fly Ash Filter: followed by sedimenttation	Laboratory study with pond effluent	Limited. economical for small facility	54
Natural sedimentation: in ponds with no algal harvesting	Field study		55, 56
Dissolved air flotation: alum or polymers	Laboratory, pilot, and field studies	Extensive, authors claimed economical use	57, 58

quiet place for oviposition. Emergent vegetation also provides a site for oviposition. Mosquito breeding is extremely heavy where overflow water is allowed to drain across a flat area before reaching a drainage ditch. The swampy areas with vegetation are ideal places for breeding by providing cover and food. Screens on effluent boxes, careful construction, and proper maintenance can eliminate all of these problems.

Certain midges breed prolifically in stabilization ponds, with adults becoming a nuisance around residential areas because of their intolerable numbers. The problem is one of nuisance rather than of health hazard, although mechanical transmission of pathogen by midges is also possible. Emergent vegetation also has significant influence on the occurrence of many aquatic insects other than mosquitoes. The most obvious is the increase in the number of beetle larvae and adults, dragonfly naiads, and Hemiptera. Near shore and shallower water where sponge-like algal mats are found are places showing dense populations of midge larvae. At a distance greater than 5 to 8 ft from shore and about 2.5 ft in depth these algal mats and the dense populations of midges associated with them are seldom found. An effective control of mosquito breeding therefore is also effective for other insects.

5. CAPITAL AND OPERATING COSTS

Estimating the installed capital costs of stabilization ponds is a complex matter, according to Parker (60). Cost estimate based on surface area or flow alone is relatively meaningless. To make a reasonable estimate, most or all of the following factors must be considered:

1. Land availability and costs.
2. Surface area.
3. Depth.
4. Configuration.
5. Terrain.
6. Dam or dike construction.
7. Volume and type of earthwork.
8. Miscellaneous construction.
9. Lining (if needed).
10. Engineering and contingency costs.

In addition, costs for transfer or interconnecting pipes and channels as well as transfer pumps and their installation for a multiple-pond system could be significant. Costs for aeration or mixing equipment and their installation, if needed, should be included. Construction and installation costs for the algal removal unit need to be considered.

The following costs apply only to the basic construction cost without or with liner of a single pond, and without considering aeration or mixing equipments. The three major cost items are liner, land value, and earthwork. Tables 8.13 and 8.14 give cost estimates for earthmoving and plastic liner cost. If specific information is lacking, land cost may be estimated as $3000/acre for isolated rural areas and $15,000/acre in general industrial zones. The overall cost estimates based on pond surface area, approximate depth and location in United States are summarized in Tables 8.15 and 8.16. Table 8.17 gives costs for some actual pond installations (60).

Table 8.13
Earthmoving costs (Updated to 2005 US Dollar Value)

Category	Average cost USD/yd^3 except as noted	Cost range USD/yd^3
Sand and sandy soil		
1. Topsoil stripping, stockpiling and respreading	2.90	—
2. Cast in place and compacted (very large volume)	0.30	0.20–0.47
3. On-site panning, placing and compacting (very large volume)	1.89	0.72–2.17
4. On-site cut and fill panning and compacting (other than very large volumes)	2.90	1.31–5.80
5. Double handling for drying, mixing or other purposes	4.35	—
6. Off-site borrowing, placing and compacting	8.24	4.35–14.49
Clay and rocky soil		
7. Scrapers and bulldozers only	5.07	4.35–5.07
8. Ripping	10.14	8.70–11.59
9. Blasting	31.89	28.98–34.79
Miscellaneous Costs		
10. Off-site supplies		
a. Clay	11.59	10.14–15.94
b. Crushed stone	46.38	—
c. Rip-rap	72.48	—
11. Finish grading (pond bottom)	0.14	—
12. Landscaping and seeding	0.87/yd^2	—

Table 8.14
Material costs for plastic liners (Updated to 2005 US Dollar Value)

	Price, USD/ft^2			
Material	1000 to 10,000 ft^2	10,000 to 100,000 ft^2	100,000 to 200,000 ft^2	Over 200,000 ft^2
Polyvinyl chlorides				
10 mil	0.34	0.31	0.30	0.30
15 mil	0.56	0.44	0.41	0.41
20 mil	0.64	0.55	0.55	0.51
30 mil	0.95	0.84	0.81	0.78
Chlorinated polyethylene				
20 mil	0.75	0.64	0.61	0.61
30 mil	1.11	0.95	0.92	0.92
30 mil	1.79	1.56	1.42	1.39
Hypalon				
30 mil	1.42	1.22	1.15	1.15
30 mil (reinforced)	2.03	1.76	1.62	1.59

Table 8.15
Unlined pond construction costs[b] (Updated to 2005 US Dollar Value)

Pond Size	Cost/acre USD	Cost/acre-ft USD
1 acre		
4 ft depth	19,700	9,900
8 ft depth	39,100	6,500[a]
20 ft depth	127,600	7,100[a]
10 acres		
4 ft depth	3,600	1,800
8 ft depth	8,600	1,500
20 ft depth	36,400	2,000[a]
50 ft depth	185,600	3,800[a]
100 acres		
4 ft depth	1,100	550
8 ft depth	2,600	440
20 ft depth	11,600	640
50 ft depth	58,700	1,200[a]
1,000 acres		
4 ft depth	—	—
8 ft depth	840	140
20 ft depth	3,700	200
50 ft depth	18,300	370

[a] Not corrected for slope effects on volume.
[b] No land value Included.

Table 8.16
Lined pond construction costs[a] (Updated to 2005 US Dollar Value)

Pond area	USD	USD
1 acre		
4 ft depth	60,900	133,400
8 ft depth	82,600	156,500
20 ft depth	177,600	263,800
10 acre		
4 ft depth	36,500	89,900
8 ft depth	42,000	95,700
20 ft depth	72,500	129,000
50 ft depth	226,100	289,900
100 acre		
4 ft depth	31,900	82,600
8 ft depth	34,800	85,500
20 ft depth	42,000	92,800
50 ft depth	91,300	145,000

[a]No land cost included.

Table 8.17
Unlined pond costs taken from actual installations[a] (Updated to 2005 US Dollar Value)

Pond size acres	Cost/acre USD
Less than 10 ft useful depth	
0.6	46,400
1	34,800
1.5	38,600
4	21,800
5	7,300
10	11,600
13	8,700
25	1,500
100	3,300
125	8,700
125	2,900
150	8,700
280	7,300
500	11,600
800	7,300
11–30 ft useful depth	
5	14,500
130	8,400
150	5,300
228	3,300
250	2,900
275	10,000
287	10,100
386	2,900
650	2,400
31–40 ft useful depth	
457	15,900
41–60 ft useful depth	
110	87,100
103	98,600
234	33,200
325	58,000

[a]No land cost included.

All cost data are updated from year 1983 to reflect year 2005 Dollar value using the Cost Index for Utilities (see Appendix A); costs were multiplied by a factor of $516.75/330.82 = 1.56$ (61). The costs in 2008 Dollar values can be obtained by multiplying the given 2005 costs by a factor of $552.16/516.75 = 1.07$.

The cost of stabilization pond operation should be much less than other biological treatment processes because of its simplicity in operational requirements. Operation and maintenance

Table 8.18
Operational costs and person-hr requirements for stabilization ponds (Updated to 2005 US Dollar Value)

Item				
Flow, MGD	0.1	0.5	1.0	2.5
Weekly person-hr requirement for operation and maintenance	7.1	12.7	16.3	22.6
Annual electricity cost, USD	5,870	14,300	21,000	34,900
Annual labor cost, USD	4,100	7,360	9,430	13,140
Total annual operation and maintenance cost, USD	1,740	4,600	7,000	12,170

costs for wastewater stabilization ponds from a report by Michel (62) are also updated to 2005 and summarized by the authors in Table 8.18.

6. DEVELOPMENTS IN PONDS APPLICATIONS

There have been very few new developments in stabilization pond processes. Much of the effort in recent years has been devoted toward understanding the mechanism of substrate removal in the complex biological community and to improving methods of algal removal for better control of effluent quality. In this section the areas of recent development in the use of stabilization ponds are discussed.

6.1. Nutrient Removal and Controlled Eutrophication

It is well-recognized by environmental engineers that nitrogen and phosphorus uptake by algae in stabilization ponds contribute significantly to nutrient removal from wastewater. Nitrogen and phosphorus are removed by amount of removal by each of these removal mechanisms is dictated by the dominant pond activity, which in turn is determined by the environmental conditions of the pond, e.g., degree of mixing, radiation availability, pH, temperature, organic loading, and so on. With a better understanding of pond activities today, attempts have been made to optimize the processes of nutrient removal in stabilization ponds (63, 64).

Both Assenzo (65) and Bush et al. (66) found significant nutrient removal from ponds, ranging from 30% to 95%. By analyzing the data of several pond operations in central Oklahoma, Assenzo reports that the optimum removal of N or P depends on the BOD/N/P loading ratio of the pond, with N expressed in total Kjeldahl-N and P in total phosphorus. The BOD/N/P ratio for optimum removal of N ranges from 78/13/1 to 16/8/1, while the same ratio for optimum removal of P ranges from 52/10/1 to 21/5/1. Compromising between the optimum ratio for N and P removal, the loadings are approximately 52 lb BOD/ac./d, 12 lb N/ac./d, and 1.6 lb P/ac./d.

To demonstrate the feasibility of a luxurious N uptake by algae in high-rate aerobic ponds, Golueke et al. (67) showed that with an initial content of 35 to 45 mg/L of NH_4^+ nitrogen in the wastewater, the additional algal yield per gram of anhydrous NH_3 nitrogen is only 0.5 g and per gram of urea nitrogen is 0.6 g. However, if the initial NH_4^+ nitrogen concentration is less

than 5 mg/L, the respective algal yields are 5 and 10 g. A high NH_4^+ nitrogen concentration in excess of 45 mg/L or NO_3^- nitrogen concentration above 35 mg/L will inhibit algal growth. The maximum algal concentration is obtained when NH_4^+ nitrogen concentration is 12 to 20 mg/L.

To take advantage of the ability of algae to remove nutrients, a controlled eutrophication technique can be applied to stabilization ponds. In essence, the technique is to regulate the assimilation of nutrients by algae in mass culture. Furthermore, the algal mass produced can be biologically removed by linking the system to a commercially valuable culture of plankton-feeding invertebrates such as oysters, clams, mussels, brine shrimp, or fish (68–73). Although the feasibility of such a controlled eutrophication system has been successfully demonstrated, the scaleup presents engineering problems. In these nutrient-algae-aquaculture systems, one should not lose sight of the primary objective of nutrient removal while mass culture of algae and balancing of the ecosystem with invertebrates are merely viable means to achieve the goal of nutrient removal. This is a very important consideration because it has been found that various requirements needed to meet the separate objectives of nutrient removal and producing algae for aquaculture are not entirely compatible. Goldman et al. (74) showed that increasingly higher removals of nitrogen through algal assimilation will occur as the dilution rate or the fraction of wastewater in the influent is decreased. On the other hand, the maximum yield of algae will occur at a relatively high dilution rate, leaving a relatively high nitrogen concentration.

To optimize nutrient removal in the nutrient-algae-aquaculture system, it is important to maintain NH_4^+ nitrogen as the dominant N form in the wastewater. In a massive algal growth system, the pH of the water will reach as high as 10.5 at times during the day. Stripping of NH_3 to the atmosphere occurs at the high values of pH typically attained in these ponds. High nitrogen removal (70% to 75%) occurs because of the combination of algal assimilation and NH_3 stripping. Therefore, for maximum nitrogen removal, nitrification of the wastewater at the treatment plant preceding the algal system must be avoided. Removal of phosphorus is also best achieved at high pH in fresh as well as in saline water environment because of precipitation. Only 25% to 50% of the phosphorus in a typical secondary effluent could be expected to be removed through incorporation into algal cells. The precipitated phosphorus, if not physically removed, could revert into solution when it comes in contact with receiving water at a lower pH.

On average, the maximum attainable yield of algae is 6 g of particulate carbon/m^2/d, or approximately 12 g dry weight of algal cells/m^2/d. Local temperature and sunlight effects will control the practical limit of algal production and in turn the yield of aquaculture product.

6.2. Integrated Anaerobic-Facultative-Aerobic Pond Systems

An integrated pond system consists of an anaerobic pond, several facultative ponds, and an aerobic pond in series. The anaerobic pond is isolated at the center of the series and is surrounded by the facultative ponds. All these ponds in turn are surrounded by a buffer pond of indefinite size and shape designed mainly for water dissipation and esthetic appeal. Effluent is drawn from near the bottom of the anaerobic pond to retain grease and heat. Moving

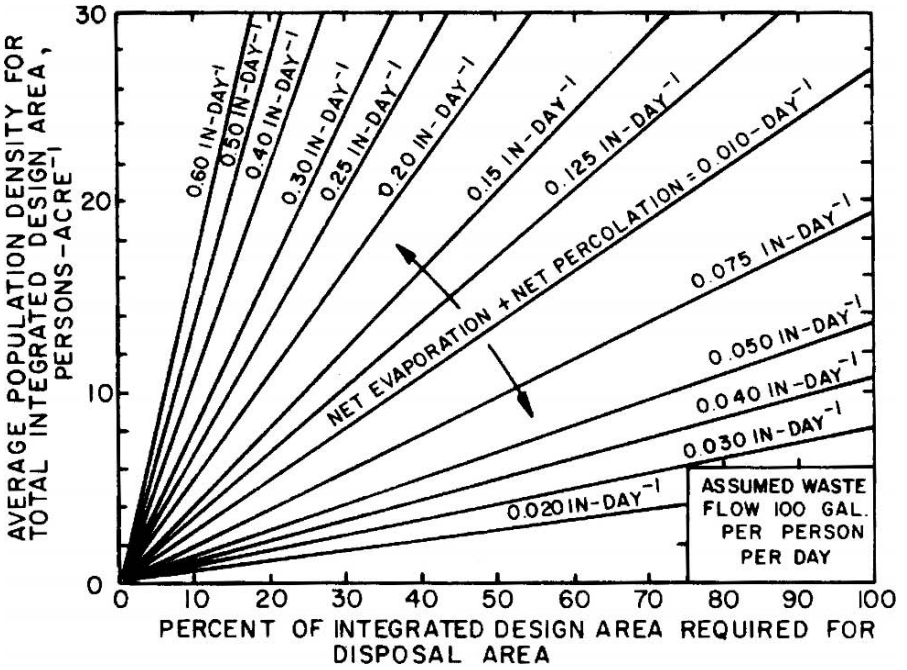

Fig. 8.5. Relationship between pond area and population density for various rates of evaporation and percolation.

through the facultative ponds in series, the effluent is retained in the buffer pond. A shallow well can be installed in the berm that separates the last facultative pond and the buffer pond. Water from the well could be used for park irrigation or to maintain the water level in the ponds.

The integrated pond system, conceptually designed by Oswald et al. (75) was to replace septic tank systems for subdivisions. A hydraulic balance had to be maintained so that the inflow is equal to the net evaporation and net percolation. There was no overflow from the system. Hydrological data on precipitation and evaporation as well as percolation rates through soil at the site were analyzed and carefully evaluated. An idealized general relationship between population density in the subdivision, wastewater flow, net evaporation plus percolation, and percent of total available area required for the system is illustrated in Figure 8.5.

The required area for the system was less than that of septic tank systems. It is an attractive solution for subdivision wastewater disposal in areas where sewerage is not yet available and will not be available within several years. In many areas septic tanks are not feasible because of experienced failures or because percolation tests show that required leaching field areas are greater than the size of existing lots. However, the system should be installed on a fixed-population basis and in areas where local surface water or ground water supply is not used for drinking water.

The buffer pond, through skillful design by landscape architects, can contribute immensely to community acceptance of the system. Operating experience at the City of Esparto and Santee, both in California, or similar ponds in series show that no sight or odor nuisance, nor hazard to health would result from using the integrated pond-park concept. Employing ponded wastewater in parks for viewing and for contact sports, including even fishing and swimming has been accepted at Santee (76).

6.3. Activated Sludge Process Integration

The activated sludge process (ASP) can be successfully integrated into the ponding system in the framework of the pond enhanced treatment and operation (PETRO) concept (77). The PETRO system, an ASP-variant, was designed to combine oxidation ponding as a low tech primary stage and a polishing facility as a secondary stage (78). The full scale system has two variants in which the secondary facility can be either a trickling filter or an activated sludge process. A series of oxidation ponds treat the bulk of organic load (up to 70%), which substantially decreases the size of the relatively high tech secondary facility.

It was demonstrated that microalgae, acting as heterotrophs in the dark, contribute to the mucilage production in the trickling filter. Biofilm slime, predominantly exopolysaccharide, was shown to be produced by the microalgae experiencing stress transfer from the mixotrophic (growth on organics in the light) conditions in the ponds to the heterotrophic conditions in the trickling filter (metabolism in the dark). Microalgae are thought to use low molecular weight organics such as amino acids, monosaccharides, VFA, etc. Microalgae appear to play a prominent part in the system and more importantly, help produce the superior quality of its final effluent.

The relative importance of the different mechanisms involved in the removal of microalgae in the trickling filter and activated sludge process varies. Removal by rotifers and protozoa plays a greater role in the trickling filter. The removal in the trickling filter is characterized by both flocculation (owing to algal and bacterial exopolysaccharide production) and degradation (through bacterial activity) of algae. In the activated sludge process, algae removal is achieved primarily through the stress-induced exopolysaccharide production (both algal and bacterial) and subsequent flocculation (embedding of the algal biomass in the sludge flocs).

It was suggested that using the PETRO ASP as secondary treatment produce better sludge flocculation. Reciprocally, better flocculation would lead to higher retention of slow-growing nitrifiers enhancing the nitrification potential of the ASP. The higher nitrification rate owing to lower heterotrophic conditions is particularly important at lower temperatures (below 15°C). At the lower substrate concentrations a nitrifying population is built up at shorter sludge ages in PETRO ASP than in a conventional ASP.

6.4. Integrated Duckweed and Stabilization Pond

Posttreatment of effluent from an up-flow anaerobic sludge blanket (UASB) reactor, which was fed with domestic wastewater, was conducted in an integrated pond system (79). The system consisted of a series of shallow duckweed and stabilization ponds. The main objective of post-treatment is removal of bacterial pathogens and further polishing of effluent quality. Rapid and efficient pathogen removal can be achieved in shallow stabilization ponds but their

effluent BOD and TSS is relatively high, owing to the presence of algae. Passing stabilization pond effluent through duckweed ponds was expected to remove algae owing to reduced light penetration. Duckweed ponds have revenue-generating potential because the produced biomass can be used as animal fodder. However, when applied separately, their pathogen removal is poor. At a pilot plant system with an overall retention time of 4.2 d consisted of 10 ponds in series, 2 duckweed ponds in first stage, 3 stabilization ponds in the second stage and 5 duckweed ponds in the third stage, rapid removal took place in the stabilization ponds. Increasing the retention time of the stabilization ponds to 3 to 4 d is suggested for consistently satisfying the WHO criterion for unlimited irrigation. A first order fecal coliform decay constant K_d was calculated for each of the three stages with the values of 0.7 to 3.2, 4.0 to 5.9 and about 1.4/d, respectively. The shading by the duckweed cover in the last stage proved to be able to remove practically all algae. Excellent effluent quality with respect to TSS (11 mg/L) and wastewater treatment for reuse in irrigation could be achieved in one simple system.

Ammonium removal was due to uptake by the duckweed plants, nitrification, sedimentation, combined volatilization of NH_3 and denitrification (80). The duckweed plants were shown to release oxygen to the pond water remained completely aerobic and could remove 95% to 99% of the influent BOD (81). The duckweed ponds can be situated in the treatment scheme before and after the algae pond treatment; in the former case, to benefit from high nutrient concentrations, and in the latter, to remove the algae from the algae pond effluent and for additional nutrient conversion. Further information on nitrogen removal, attached growth waste stabilization ponds and algal ponds can be found in refs. (82, 83).

6.5. Deep Self-regeneration and Anoxic Waste Stabilization Ponds

Waste stabilization ponds are becoming the most popular forms of secondary treatment owing to their low cost of construction, operation and maintenance. However, as stated above, the presence of algae in the effluent makes its quality unacceptable in terms of total suspended solids and biochemical oxygen demand. Stringent environmental regulations concerning effluent discharges make it necessary to remove algae from pond effluent.

Anoxic waste stabilization ponds operate in the so called “gray area” of organic loading, which lies between fully anaerobic and facultative conditions. In anoxic waste stabilization ponds, a removal efficiency of suspended solids of 60% to 80% and biochemical oxygen demand and chemical oxygen demand (COD) of 70% to 80%, algae of 76% to 98% can be achieved. Regarding algal species in anoxic pond effluent, the motile flagellate algae (*Euglena and Chlamydomonas*) were the only species found to exist.

In pond systems, it is known that algae stimulate bacteria, and bacteria stimulate algae. In this way, the chemical units that comprised the organic waste eventually become incorporated into the algae as stable organic components of living cells. The death and decomposition of large numbers of algae lead to undesirable conditions similar to those caused by the original wastewater in the pond. Consequently, it needs to be avoided to keep the trophic structure well balanced for optimum performance. Deep treatment ponds offer theoretical advantages over conventional stabilization ponds, which have greater surfaces and are shallower. They occupy less surface area and, in arid and semi-arid areas, summer evaporation losses are reduced.

Deep ponds could become an inexpensive, simple way to improve the sanitary and chemical quality of water and, at the same time, be used as reservoirs for agricultural irrigation. Arauzo et al. (84) carried out investigation of experimental deep self-regeneration pond involving the in-depth examination of the role of algae in a deep urban wastewater self-regeneration pond. Evaluation was made of the phytoplankton community, the dynamics of its populations and its relationship with the performance and the trophic structure of the pond. The aim of their design was to improve the quality of wastewater effluent for its reuse.

Despite their fertilizing value and protein content, algae are required to be removed owing to their deleterious effect on the receiving environment. Accordingly, if pond treatment is to remove BOD from the influent, then algae should be removed from the effluent (85). The US EPA considered the removal of algae from ponds effluent as a condition for successful operation and production of good quality effluent (86). Different treatment processes were applied to upgrade effluents from waste stabilization ponds.

Filtration is commonly employed to upgrade waste stabilization pond effluent (87, 88). Disadvantages of slow sand filtration include the deep penetration of small algal species. If these fine particles are present in large amounts, cleaning of the filter cannot be achieved by normal scraping methods. Furthermore, anaerobic conditions may develop in these filters, especially during darkness or at low temperatures, owing to decomposition of algae which will result in production of soluble BOD, deficiency of oxygen, production of odor and release of nutrients. Owing to the increasingly restrictive legal requirements, the rising cost of liquid waste disposal, and the growing need for innovative sources of water supply, liquid waste reduction and recovery of valuable products, cross-flow membrane separation processes are gaining a considerable prominence in many sectors of the industry (89).

Cross-flow filtration is a process in which the formation of a filter cake is either limited or, under certain conditions, almost completely suppressed by a flow of the suspension parallel to the filtration surface and, because this system is pressurized, water is forced through the filter. In contrast, in the case of conventional cake filtration, the suspension flows at right angles to the filter medium under the applied pressure. The particles are retained by the filter medium while the liquid flows through the filter cake and through the medium. It is clear that, in the case of conventional filtration, clogging will occur in a short time owing to the build-up of filtered cake, while in cross-flow filtration particles deposited on the filter medium are swept away by the cross-flow velocity actions.

6.6. Algae and Phosphorus Removal by Induced Air Flotation

Algae have many of the properties of fine near-colloidal particles, ranging in size from the submicron to about 20 μm in diameter or length. They can be flocculated and floated using induced air flotation system. Tests have been done on water from a wastewater maturation pond (90). The flocculant of iron chloride added at an average dose of 10 mg/L. It seems to be very effective in the removal not only of the algae, but also of the phosphorus compounds. At a pH of 9 in water, the iron salt hydrolyses to form mainly ferric hydroxide, which has excellent flocculating properties. The soluble phosphates in the water reacted with the iron salts to form an insoluble precipitate, which was flocculated with ferric hydroxide produced by a stoichiometric excess of iron chloride (87, 88). The removal efficiencies of algae cell

count and total phosphorus were 98.8% and 94.2% respectively. The excess of iron chloride addition could be simultaneously removed with flotation.

6.7. Combination with Constructed Wetlands

Recently, the use of constructed wetlands (91) has been receiving more attention for various types of wastewater treatment alternatives (92). The inclusion of constructed wetland technology as a component of an overall wastewater management system has the potential to help address the issues of agricultural wastewater treatment and the protection of water quality. An integrated wastewater treatment facility, consisting of upper (solids separators, anaerobic lagoons, and aerobic ponds) and lower (wetland cells) subsystems, has been built to replace the lagoon (93). The demonstration of the system for dairy treatment is that the collection sump of the new waste treatment facility collects all dairy wastewater outflows. Wastewater is then pumped to solids separators, and flows by gravity to anaerobic ponds and aerobic ponds. The upper subsystem is expected to treat the water sufficiently so that the wetland cells may achieve further pollutant reductions. The lower subsystem, comprised of 8 surface wetland cells with approximate surface of $5{,}000\,m^2$, receives outflow from the ponds. The cells were planted with cattail, soft-stem bulrush, and reed. After treatment is completed through the lagoons and ponds followed by the wetland cells, the wastewater can be reused to flush barns or to irrigate crops.

An integrated aquaculture–wetland ecosystem using tertiary-treated municipal wastewater in Los Angeles County, California was proposed for food production and nitrogen removal (94, 95). The ecosystem connected polyculture aquaculture ponds with in-pond aquatic plant systems, a solar energy aeration system and an artificial wetland. Ponds were stocked with hybrid tilapia, common carp, mosquitofish, and red swamp crayfish, and were flushed weekly with new wastewater. The concept of using tertiary-treated wastewater for aquatic food production may be attractive in the periurban areas of many meagcities, both for fish markets and to stem the growing discharges of wastewater that are causing coastal pollution.

An index of biotic integrity based on crayfish, fish, and amphibian assemblages was developed to assess vernal ponds and palustrine wetland habitats along the southern shore of Lake Michigan (96). The other experimental study of constructed wetlands included multi-stage systems (97). Pretreated water from an anaerobic stabilization pond and treated water from the last pond of a lagoon system were used to test the system's suitability as a complementary system for removing nitrogen and phosphorus (98).

Seepage losses from constructed wetlands, wildlife refuges, wastewater lagoons, runoff collection ponds, and other engineered surface impoundments of water or aquatic ecosystems in areas with deep groundwater levels can be minimized with artificial liners (plastic, compacted earth, etc.) or with natural processes like sediment accumulation and microbiological and chemical processes (99). Sediment accumulation can be increased by deliberately adding soil slurries or muddy (turbid) water. When these are applied to the water surface, settling of the particles to the bottom then creates a graded sediment layer with the coarsest particles on the bottom and the finest particles on top. Column studies in the laboratory showed that, for a given amount of soil added, such a graded layer gives more seepage control than a compacted earth liner. Also, split slurry applications to create a layered lining gave more seepage control

than when the same total amount of soil was given in one single slurry application. Subsequent addition of sodium carbonate to the water reduced seepage even more, giving a total seepage reduction in the columns from 1000 to 0.2 cm/day. The study also indicated the importance of avoiding turbid inflows into infiltration basins for groundwater recharge, as well as soil erosion within the basins themselves, if infiltration rates need to be maximized.

6.8. Synopsis of Major Developments

Waste stabilization ponds design practice today is much different from what used to be 20 years ago. There is now a new generation of waste stabilization ponds that are smaller, more efficient, more reliable, with less or no odor problems, and which release effluents of better quality than the previous generation. In summary, numerous different types of ponds have been developed during the years: anaerobic ponds (open or covered); aerobic ponds (maturation or polishing); facultative ponds; ponds with or without recirculation; high rate stabilization ponds (HRSP); partially mixed and bi-dimensional plug-flow ponds; ponds for both municipal and industrial effluents, etc. (100). To locate names and web sites of manufacturers and products for stabilization ponds and lagoons consult ref. (101).

The main new concepts and tools available for the designers of modern waste stabilization ponds are in the fields of hydraulics, kinetics, out-of-the-pond elements and design of the wastewater treatment series (100):

(a) Hydraulics:

1. Actual hydraulic pattern (tracer studies) and development of tools to improve it.
2. Pond shape.
3. Wind effect on wastewater circulation within the pond.
4. Stratification control.
5. Nonsteady-state approach to reactor analysis (102).
6. Use of both horizontal and vertical baffles to force plug-flow conditions (103).
7. Sequential batch operation.

(b) Kinetics:

1. Pond acceleration by low-energy mixing, aeration or re-circulation.
2. Pond acceleration by addition of fixed biomass.
3. Combination of ponds with more intensive treatment units.

(c) Out-of-the-pond elements:

1. Different types of pond covers and devices to avoid smell release.
2. Intensive anaerobic pretreatment to reduce pond size.
3. Posttreatment for algae removal (rock filters, constructed wetlands) and other requirements.

(d) Design of the wastewater treatment series:

1. Ponds of different sizes instead of modularization.
2. Use of wastewater storage reservoirs for irrigation, controlled discharge and other purposes.

7. EXAMPLES OF PROCESS DESIGN

In this section, four examples of process design are given. The first one illustrates an aerated pond design. With consideration of the odor problem caused by the presence of H_2S in pond water, the second example is used to show how the problem can be averted

in a facultative pond design. Still another approach to design involving heat preservation is illustrated in the third example. Finally, the rational design of a high-rate aerobic pond is given.

7.1. Example 1

Given:

Influent soluble $BOD_5 = 150\,mg/L$
Suspended solids negligible
Flow = 1 MGD
Flow temperature = 15°C or 59°F
Atmospheric temperature, summer = 30°C or 86°F, winter = 10°C or 50°F
First order BOD_5 removal rate: $k = 2.51/d$ at 20°C for Equation 14 and 0.25 1/d for Equation 18
Mechanical aerator rated at 3.0 lb O_2/hp/h by the manufacturer

Required:

Design an aerated lagoon to provide an effluent soluble BOD_5 of 15 mg/L.

Solution: Complete-Mix, Aerated Pond

1) Estimate the summer and winter pond temperature, using Equation 36, rewritten as:

$$T_w = \frac{QT_i + AfT_a}{Af + Q}$$

Select 1.0 acre or $43{,}560\,ft^2$ for pond surface area

$$\text{Summer: } T_w = \frac{1.0(59) + 43{,}560(12 \times 10^{-6})(86)}{43{,}560(12 \times 10^{-6}) + 1.0} = 68.4°F \text{ or } 20.2°C$$

$$\text{Winter: } T_w = \frac{1.0(59) + 43{,}560(12 \times 10^{-6})(50)}{43{,}560(12 \times 10^{-6}) + 1.0} = 56°F \text{ or } 13°C$$

2) Estimate the detention time using Equation 14 and select from Table 8.7 a temperature coefficient. $\theta = 1.04$

$$\text{Summer: } k_{20.2°C} = 2.5(1.04)^{20.2-20} = 2.6$$

$$\frac{15}{150} = \frac{1}{1 + 2.6(t)}$$

$$t = 3.46\,d$$

$$\text{Winter: } k_{13.3°C} = 2.5(1.04)^{13.3-20} = 2.13$$

$$\frac{15}{150} = \frac{1}{1 + 2.13(t)}$$

$$t = 4.23\,d$$

Using 4.23 d, calculate effluent BOD_5 for summer time

$$\frac{S}{150} = \frac{1}{1 + 2.6(4.23)}$$

$$S = 12.5\,\text{mg/L BOD}_5$$

3) Depth of pond

$$D = \frac{Qt}{A} = 13\,\text{ft}$$

4) Estimate the biological solid production, using Equation 6 in Chapter 6. Activated sludge processes

$$X = \frac{Y(S_o - S)}{1 + bt}$$

Summer conditions are used in this calculation because higher BOD removal is effected and consequently more solid production and more oxygen consumption are expected.

Also assume: $Y = 0.65$

$K_s = 100\,\text{mg/L}$, and

$b = 0.11/\text{d}$

$$X = \frac{0.65(150 - 12.5)}{1 + 0.1(4.23)} = 63\,\text{mg/L } VSS$$

5) Estimate the oxygen requirement following Example 2 from the Chapter 6. Activated sludge processes

$$O_2\text{lb/d} = \frac{(150 - 12.5)(8.34)(1)}{0.68} - 1.42(63)(8.34)(1) = 1{,}680 - 750 = 930\,\text{lb/d}$$

6) Estimate oxygen transfer rate under field conditions assuming that the field condition is elevation at 1,000 ft. From Table 8.10, $C_s = 8.8\,\text{mg/L}$, selected $\alpha = 0.85$, $\beta = 0.95$, and C_r at 1.0 mg/L most be maintained. From Equation (38)

$$N = (3.0)\frac{[(8.8)(0.95) - 1.0](1.024)^{20.2-20}(0.85)}{9.17} = 2.09\,\text{lb O}_2\text{/hp/h}$$

7) Estimate the horsepower requirement for the mechanical aerator,

$$\text{hp} = \frac{930}{(2.09)(24)} = 18.6$$

8) Estimate the horsepower requirement for complete-mix in pond, using a power requirement of $0.5\,\text{hp}/1000\,\text{ft}^3$

$$\text{hp} = \frac{(0.5)(13)(43{,}560)}{1{,}000} = 278$$

Thus, one should use eight 35-hp aerators.

Solution 2: Aerated Pond, Mixing Not Complete (Aerobic–Anaerobic Pond)

Referring to Figure 8.4, the value of the dispersion factor, d, ranges from zero for an ideal plug flow regime to infinity for complete-mix flow. Stabilization ponds designed as complete-mix systems would have d values from 4.0 to ∞. Most ponds, however, are somewhere within the range from 0.1 to 2.0. In this design problem, the d value is chosen as 1.0. Therefore from Figure 8.4, kt = 5.0 for reducing BOD from 150 to 15 mg/L.

1) Determine the detention time under winter conditions because the BOD_5 removal rate is slower and the larger surface area requirement of the pond controls the design.
   Select temperature coefficient $\theta = 1.05$ from Table 8.7 and assume the pond temperature in winter is 11°C for the first trial,

$$k_{11} = 0.25(1.05)^{11-20} = 0.155$$
$$0.155\,t = 5$$
$$t = 32.2\,d$$

   Use a pond depth of 5.0 ft (see Table 8.8)
2) Estimate pond surface area A,

$$A = \frac{10^6 \times 32.2}{5 \times 43{,}560 \times 7.48} = 19.3 \text{ ac.}$$

3) Check with Equation 36 on the pond temperature with the surface area obtained from the previous step,

$$T_w = \frac{1.0(59) + 19.3(43{,}560)(12 \times 10^{-6})(50)}{19.3(43{,}560)(12 \times 10^{-6}) + 1.0} = 51°\text{F or } 10.6°\text{C}$$

   Many trial-and-error steps will be required if the calculated T_w does not check closely with the assumed pond temperature.
4) Determine the horsepower requirement for mechanical aerators. The field O_2 transfer rate at 2.09 lb O_2/hp/hr has been calculated in Solution 1. Use 1.5 lb O_2/ lb BOD removal requirement,

$$\text{hp} = \frac{1.5(150)(8.34)(1)}{2.09(24)} = 37.5\,\text{hp}$$

Considering the pond area and horsepower requirement, an arrangement of 4 ponds of 5 ac. each and an aerator of 10 hp in each pond would seem desirable.

Comparing solutions 1 and 2, it becomes evident that a greater depth with much smaller surface area must be used for the complete-mix aerated pond. The horsepower requirement is significantly greater, primarily to satisfy the complete-mix requirement. For the same degree of treatment, the aerobic-anaerobic ponds have relatively much larger surface areas, shallower-depths, and much lower horsepower requirements.

7.2. Example 2

Given: Same as in Example 1, but influent wastewater also contains 800 mg/L SO_4^{2-}

Required: Design a facultative pond to produce the same effluent soluble BOD_5 of 15 mg/L and completely eliminate the odor problem. The critical level of $SO_4{}^{2-}$ in the pond for algae toxicity and odor occurrence is 4 mg/L (no H_2S production is expected in the aerated ponds).

Solution:

1) First design the facultative pond without consideration of SO_4^{2-} concentration.
Select from Table 8.8 an organic loading of 75 lb BOD_5/ac./d,

$$\text{Pond area} = \frac{150(8.34)(1)}{75} = 16.7 \text{ ac.}$$

Choose a pond depth of 6 ft

$$\text{Detention time } t = \frac{V}{Q} = \frac{43{,}560(16.7)(6)}{1.547(24)(3{,}600)} = 32.8 \text{ d}$$

2) Estimate the sulfide concentration based on Equation 41

$$\begin{aligned} S^{2-} &= (SO_4^{2-})[0.0001185(\text{BOD Surface load}) - 0.001655(\text{t}) + 0.0553] \\ &= 800[0.0001185(75) - 0.001655(32.8) + 0.0553] = 800[0.01] = 8 \text{ mg/L} \end{aligned}$$

3) Redesign the pond to remove 4 mg/L of S_2^- in the wastewater.

$$\begin{aligned} S^{2-}/SO_4^{2-} &= \frac{4 \text{ mg/L}}{800 \text{ mg/L}} = 0.005 \\ &= 0.0001185(\text{BOD}) - 0.001655(\text{t}) + 0.0553 \end{aligned}$$

One alternative is to keep the BOD surface load constant, and increase the detention time, i.e., increase the depth of the pond.

$$\text{Detention time} = \frac{1}{0.001655}[0.0001185(75) + 0.0553 - 0.005] = 35.8 \text{ d}$$

$$\text{Pond depth} = 6.55 \text{ ft}$$

Another alternative is to keep the pond depth at 6 ft and change the detention time and surface area.

$$\text{Detention time} = \frac{1}{0.001655}[0.0001185(\text{BOD}) + 0.0553 - 0.005]$$

Where

$$\begin{aligned} \text{BOD surface load} &= \frac{150(8.34)(1)}{\text{Pond surface area in acre}} \\ &= \frac{150(8.34)(1)}{(t)(1.547)(24)(3{,}600)/(6)(43{,}560)} \\ t &= \frac{1}{0.001655}\left[0.0001185\left(\frac{2{,}450}{t}\right) + 0.0553 - 0.005\right. \end{aligned}$$

Solving for t:

$$t = 35.5\,\text{d}$$

$$\text{Surface area A} = \frac{35.5(3{,}600)(24)(1.547)}{43{,}560(6)} = 18.1\,\text{ac.}$$

which is an increase of 1.4 ac. From the original pond area of 16.7 ac.

7.3. Example 3

Given: Same as in Example 1.

$$\text{BOD removal rate } k = 0.251/\text{d at } 20^\circ\text{C}$$

Required: Design a stabilization pond to minimize heat loss and to provide an effluent soluble BOD_5 of 15 mg/L.

Solution:

1) Select a temperature coefficient $\theta = 1.06$ and assume the pond wastewater temperature $T_w = 21^\circ$C, or 69.8°F in summertime,

$$k_{21^\circ\text{C}} = 0.25(1.06)^{21-20} = 0.265\,\text{d}^{-1}$$

Substitute into $0.9 = 1 - e^{-kt}$

and solve for $t = 8.7\,\text{d}$

2) Rewrite Equation (36) to yield

$$\frac{1}{134{,}000}\frac{D}{t} = \frac{f(T_w - T_a)}{T_i - T_w}$$

$$\frac{D}{t} = \frac{1.6(T_w - T_a)}{(T_i - T_w)}$$

Choose a depth $D = 20\,\text{ft}$ and substitute into the equation to solve for T_w

$$\frac{20}{8.7} = \frac{1.6(T_w - 86)}{(59 - T_w)}$$

$T_w = 69.8^\circ$F or 21°C; same as assumed value

3) Repeat steps 1 and 2 by assuming pond temperature $T_w = 13.5^\circ$C or 56.3°F in the winter,

$$k_{13.5} = 0.25(1.06)^{13.5-20} = 0.172\,1/\text{d}$$

Substitute into $0.9 = 1 - e^{-kt}$

and solve for $t = 13.4\,\text{d}$

Choose $D = 40\,\text{ft}$.

$$\frac{740}{13.4} = \frac{1.6(T_w - 50)}{(59 - T_w)}$$

$T_w = 56.1^\circ$F; close to assumed value.

The pond depth is 40-ft with a detention time of 13.4 d and a surface area

$$A = \frac{13.4(24)(3{,}600)(1.547)}{40(43{,}560)} = 1.03\,\text{ac.}$$

Discussion:

Comparing with solution 1 of Example 1, the surface area requirement is practically the same, but the pond depth increases from 13 to 40 ft for maximum heat conservation. The significant difference in depth reflects the different temperature coefficient θ chosen for the two solutions. Anaerobic fermentation is likely to occur in deep ponds and the temperature effect on the BOD removal rate is more pronounced.

7.4. Example 4

Given: Same as in Example 1

In addition, the stabilization pond is located 35°N latitude at an elevation of 50 ft. The average winter condition represented by December provides a ratio of 0.6 for the total hours of sunlight to total possible hours. A photosynthetic energy conversion efficiency of 0.04 and algal cells composition of $C_7H_{8.1}O_{2.5}N$ are assumed.

Required: Design a high-rate aerobic pond to provide an effluent of 15 mg/L soluble BOD_5.

Solution:

Because high-rate aerobic ponds require shallow depth and large surface area, a winter temperature of 10.5°C is assumed for the pond wastewater. The validity of this assumption will be checked later.

1) Estimate the oxygen requirement to satisfy the BOD removal. First convert all BOD_5 to ultimate BOD.

$$\text{Ultimate BOD of influent } L = 150/[1 - e^{-0.25(5)}] = 210\,\text{mg/L}$$

$$\text{Ultimate BOD of effluent } L = 15/[1 - e^{-0.25(5)}] = 21\,\text{mg/L}$$

$$\text{Ultimate BOD removal} = 189\,\text{mg/L}$$

$$\text{O}_2 \text{ requirement } W_{O_2} = 189(8.34)(1)(454) = 7.15 \times 10^5\,\text{g/d}$$

2) Oxygenation factor $p = 1.67\,\text{g O}_2/\text{g}$ of algal cell synthesized, as given in Section 2, System Variables and Control in this chapter.
3) Estimate the available solar radiation.
From Table 8.5, for December and at 35°N latitude, $S_{max} = 96$, $S_{min} = 47$, corrected for elevation 50 ft and cloudiness of 0.6,

$$S = [47 + 0.6(96 - 47)](1 + 0.01 \times 5) = 80.2\,\text{cal/m}^2\text{/d}$$

4) Estimate the unit heat of combustion, h. For the algal cell formula of $C_7H_{8.1}O_{2.5}N$,

$$C = 57.5\%$$
$$H = 5.5\%$$
$$O = 27.5\%$$
$$N = 9.5\%$$

Substitute into Equation 28,

$$h = 400 + 127\frac{100[2.66(57.5) + 7.94(5.5) - 27.5}{398.9}$$
$$= 5{,}780\,\text{cal/g}$$

5) Estimate the surface area of pond by balancing oxygen supply and oxygen requirement,

$$\text{Rewrite Equation 27}: W_a = \frac{ESA}{h}$$

Combine with Equation 29: $W_{O_2} = pW_a$

$$A = \frac{hW_{O_2}}{pES} = \frac{5{,}780(7.15 \times 10^5)}{1.67(0.04)(80.2)}$$

6) Estimate detention time, with

$$k = 0.25(1.03)^{10.5-20} = 0.189\ 1/\text{d}$$
$$0.9 = 1 - e^{-0.189t}$$
$$t = 12.18\,\text{d}$$

7) Estimate pond depth,

$$D = \frac{12.184(24)(3600)(1.547)}{19(43{,}560)}$$
$$= 1.96\,\text{ft.}$$

8) Check temperature of wastewater in pond, use Equation 36,

$$T_w = \frac{QT_i + AfT_a}{Af + Q}$$
$$= \frac{1.0(59) + 19(43{,}560)(12 \times 10^{-6})(50)}{19(43{,}560)(12 \times 10^{-6}) + 1.0}$$
$$= 50.8^\circ\text{F or } 10.4^\circ\text{C, close to assumed value is acceptable.}$$

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NOMENCLATURE

a, b, c, d, a', b', c', and d' = coefficients for predicting the sulfide concentrations in facultative ponds
$a = \sqrt{1 + 4\text{ktd}}$
A = surface area of pond, cm^2
A = pond surface area, ft^2
C = concentration at any time
C = a constant characteristic of the pond
C_0 = initial O_2 concentration
C_r = oxygen concentration to be maintained in the wastewater
Cs = oxygen saturation in tap water at a given temperature and altitude
C_s = saturation concentration
d = depth of the pond, cm
d = diffusivity constant or dispersion number, dimensionless
D = axial dispersion coefficient, ft^2/hr
D = circulation depth or depth of frictional resistance, ft.
E = photosynthetic efficiency
E = evaporation, in/d
f = pond proportionality factor
F = fetch, miles
F = oxygenation factor
x = mean temporal velocity gradient, s^{-1}
h = unit heat of combustion, cal/g
H = total heat loss
H_e = heat loss owing to evaporation
H_c = heat loss owing to convection
H_r = heat loss owing to radiation
H_v = latent heat of vaporization
k = substrate removal or reaction rate constant, 1/d
K_2 = rate constant
K_d = first order fecal coliform decay constant
K_L = oxygen transfer coefficient, m/s
L = characteristic length or travel path of a typical particle in the reactor, ft
L = organic load, lb/ac./d
$m = 1/2$, 1, and 2 for U = small, medium, and large wind speed, respectively
n = number of ponds in series
N = oxygen transfer rate at field condition, lb O_2/hp/hr
N_o = standard oxygen transfer rate, lb O_2/hp/hr
p = oxygenation factor
p = a function of latitude and lies between 1×10^{-2} and 5×10^{-2}
P_a = barometric pressure, in of mercury
Q = pond flow rate, MGD

Q = flow rate
r = recycle ratio = R/Q
R = degree of reduction of cellular organic material
R = recycled flow rate
S = effluent substrate concentration from the nth pond
S = solar radiation in Langleys, cal/cm^2/d
S = visible solar radiation energy
S = substance concentration, mg/L
S_d = substrate concentration in the diluted wastewater entering the first pond
S_i, k_i, v_i and t_i = parameters for pond i
S_o = influent substrate concentration
S_o = influent substrate concentration, mg/L
t = detention time, d
t = time
T_a = air temperature, °F
T_i = influent temperature, °F
T_w = pond water temperature, °F
U = wind speed, m/s
U = fluid velocity, ft/h
V = vapor pressure at the water temperature, in of mercury
V = liquid volume
V_a = vapor pressure in atmosphere, in. Hg
V_0 = surface-current velocity
V_0/W = a term called proportional surface velocity of the water
V_w = vapor pressure at water surface, in. Hg
W = wind speed, miles/hr
W_a = net weight of algal cells synthesized daily, g/d
W_{O_2} = amount of oxygen yield, g/d or lb O_2/ac./d.
β = oxygen saturation coefficient in wastewater
α = oxygen transfer rate correction factor for wastewater
θ = temperature coefficient
μ = absolute viscosity
v = vapor pressure at dewpoint temperature of the atmosphere, in of mercury
σ_t = standard deviation
σ = Variance

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APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers (61)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

9
Trickling Filters

Lawrence K. Wang, Zucheng Wu, and Nazih K. Shammas

CONTENTS

Abstract Trickling filter consists of a fixed biological bed of rock media or plastic media on which wastewater is applied for aerobic biological treatment. Biological slimes form on the media which assimilate and oxidize substances in the wastewater. This chapter introduces the historical development, recent advances, principles, applicability, microbiology, ecology, models, types, design criteria, and case studies of this attached-growth trickling filter process.

Key Words Attached-growth biological system • trickling filter • theory • microbiology • ecology • design • case studies • low-rate trickling filter • high-rate trickling filter • Accelo-filter • roughing filter • design models.

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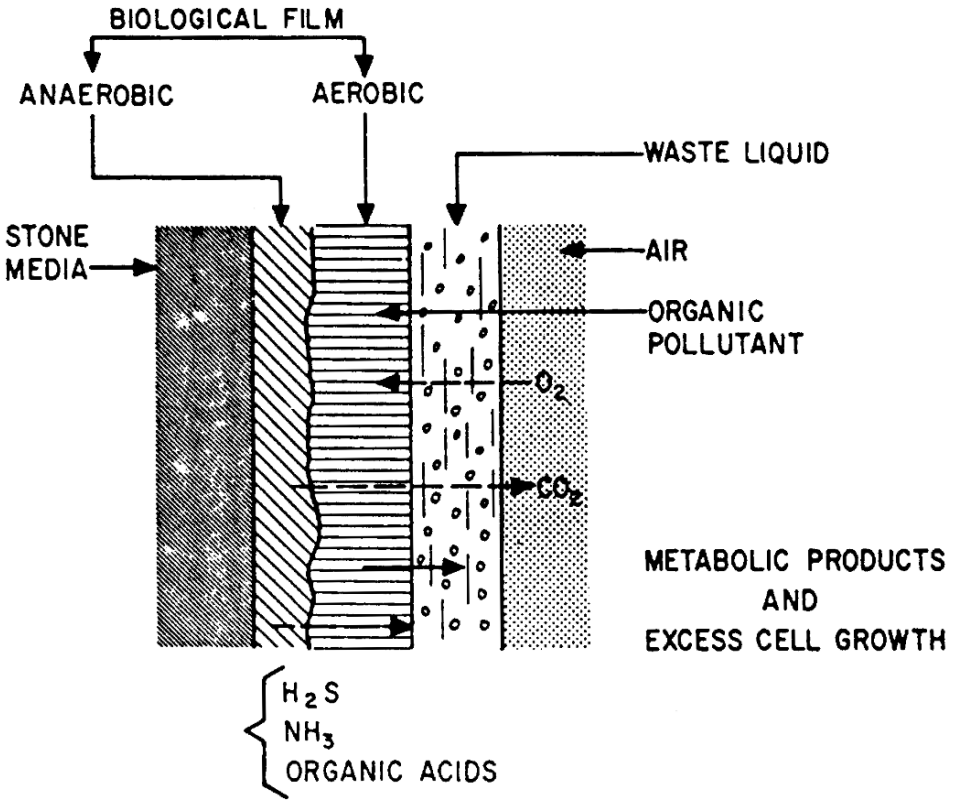


Fig. 9.1. Schematic diagram of trickling filter process (*Sources:* US EPA).

1. INTRODUCTION

1.1. Process Description of Attached Growth Systems

A contact bed, contact aerator, trickling filter, rotating discs, or other attached growth systems consist of a bed of coarse contact media such as crushed traprock, granite, limestone, clinkers, wood slats, plastic tubes, corrugated plastic sections, hard coal, or other material over which wastewater is distributed or contacted (1–9). Wastewater flows over the contact media on which a biological slime layer (i. e., zoogleal slime) develops. Dissolved organic pollutants in the wastewater are transported into the slime layer, where biological oxidation takes place. Organic pollutants are removed by the biological slime film, which consists of various microorganisms, as shown in Figure 9.1. In the outer portions of the film, organic pollutants ($C_aH_bO_cN_dP_eS_f$) are degraded by aerobic and facultative bacteria under aerobic conditions according to a biochemical reaction approximately expressed by Equation (1).

$$4C_aH_bO_cN_dP_eS_f + (4a + b - 2c - 3d + 5e + 6f)O_2 \rightarrow 4a\,CO_2 + (2b - 6d - 6e - 4f)H_2O + 4d\,NH_3 + 4e\,PO_4^{3-} + 4f\,SO_4^{2-} + (12e + 8f)H^+ \quad (1)$$

in which the end product NH_3 can be further oxidized to NO_2^- and/or NO_3^-. The microorganisms in the biological slime film grow following Equation (2) and/or Equation (3) (10–12):

$$NH_4^+ + 4\ CO_2 + HCO_3^- + H_2O \rightarrow C_5H_7O_2N + 5\ O_2 \quad (2)$$

$$C_{10}H_{19}O_3N + 1.5\ NH_3 + 2.5\ CO_2 \rightarrow 2.5\ C_5H_7O_2N + 3\ H_2O \quad (3)$$

in which $C_5H_7O_2N$ is the empirical formula of bacterial cell and $C_{10}H_{19}O_3N$ is the empirical composition of domestic wastewater (13). When the thickness of the biological film increases to a degree that the diffused oxygen is almost totally consumed before it can reach the full depth of the film, an anaerobic (or anoxic) film is established near the surface of the contact media as shown in Figure 9.1. Equation (4) describes the anaerobic biological reaction performed by facultative and anaerobic bacteria assuming all intermediate organic acids are degraded (14).

$$8\ C_aH_bO_cN_dS_e + (8a - 2b - 4c + 6d + 4e)H_2O \rightarrow (4a - b + 2c + 3d + 2e)CO_2 + 8d\ NH_3 + (4a + b - 2c - 3d - 2e)CH_4 + 8e\ H_2S \quad (4)$$

where $C_aH_bO_cN_dS_e$ is the general formula of the dissolved organic substrate penetrating the full depth of the biological film.

Because the biological slime film further increases in thickness to a degree that both dissolved organic substrate and oxygen are almost consumed before they can reach the anoxic zone, no external organic source is available for cell carbon. The facultative and anaerobic bacteria near the media face enter into a phase of "endogenous nitrate respiration" (10, 12):

$$C_5H_7O_2N + 4\ NaNO_3 + H_2O \rightarrow 4\ NaHCO_3 + NH_4HCO_3 + 2\ N_2 \quad (5)$$

in which $NaNO_3$ represents the nitrate source penetrating the full depth of the biological film. If NO_3^- is not available, the bacteria will simply decay under anaerobic or anoxic conditions:

$$2\ C_5H_7O_2N + 6\ H_2O \rightarrow 5\ CO_2 + 2\ NH_3 + 5\ CH_4 \quad (6)$$

In either case, the microorganisms lose their ability to cling to the media surface. The wastewater then washes the old biological slime layer off the contact media and a new slime layer will start to grow.

Some times the diffused oxygen can reach the microorganisms near the media face, but there is no external carbonaceous source available for cell assimilation because of either low organic loading or thick slime layer. The microorganisms near the media face will enter into an "endogeneous oxygen respiration" phase (10, 12, 15, 16):

$$C_5H_7O_2N + 5\ O_2 \rightarrow 4\ CO_2 + NH_4HCO_3 + H_2O \quad \text{or} \rightarrow 5\ CO_2 + NH_3 + 2H_2O \quad (7)$$

Similarly, the microorganisms will lose their ability to cling to the media surface and the old slime layer will be washed off the media by the shear force of the waste water flow.

The phenomenon of losing the biological slime layer in a contact bed or a similar microbial slime system is called "sloughing" and is a function of the organic and hydraulic loading on

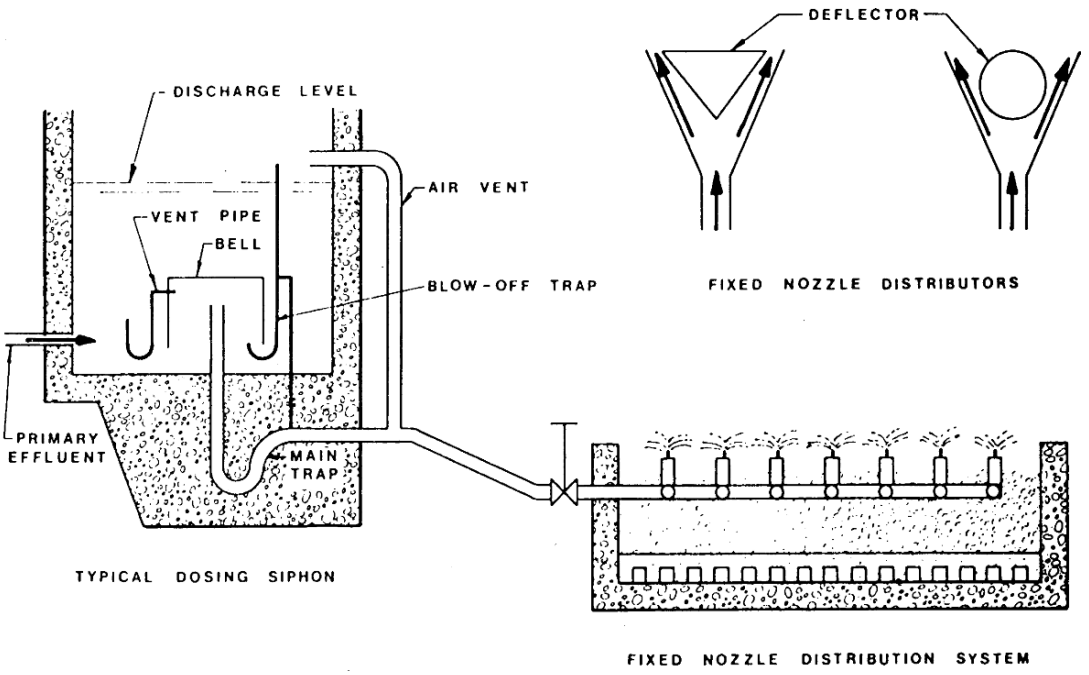


Fig. 9.2. Fixed-nozzle distribution system (*Source:* US EPA).

the process bed. Therefore, although nonsettleable solids and soluble organic pollutants are removed by an adsorption-oxidation phenomenon occurring at the biofilm-wastewater interface, the effluent from the process bed usually contains solids that are generated by periodic or continuous sloughing of the biological slime film from the process bed medium. The effluent is then subjected to a further solid – liquid separation process such as clarification. Clarification can be accomplished by either conventional sedimentation or innovative dissolved air flotation (17).

1.2. Historical Development and Applicability of Attached Growth Systems

Various contact beds and trickling filters employed for the transfer of dissolved organic matter and fine suspended solids from settled wastewater to contact surfaces have been developed (18–43). One of the artificial wastewater treatment processes used in the eighteenth century was the double-contact bed with dosing and draining siphons. The first installation of trickling filter with distribution by spray nozzles was reported to be at Lawrence Experiment Station in Massachusetts in 1891 (20). Figure 9.2 shows an early type of trickling filter with fixed nozzles and automatic dosing tank. Figure 9.3 shows a trickling filter with rotary distributor.

Biological filtration was sufficiently popular that in 1940 about 60 % of all wastewater treatment plants, which provided secondary treatment in the United States, used trickling

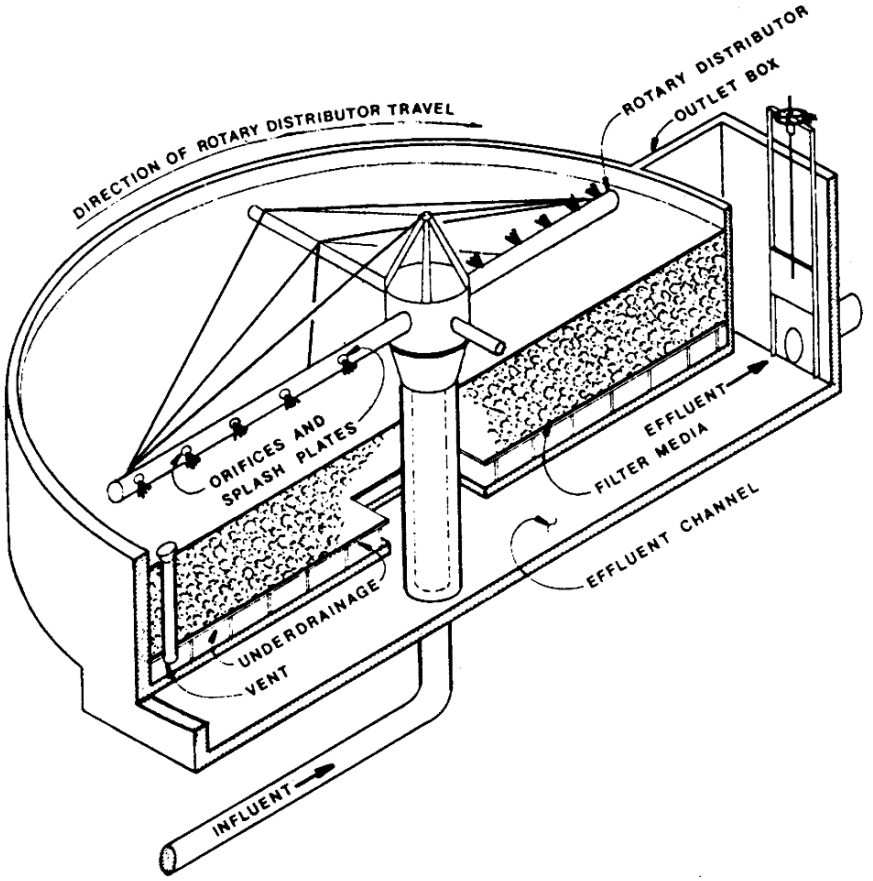


Fig. 9.3. Typical trickling filter in cross-section (*Source:* US EPA).

filters. It was in 1946 (26) that the National Research Council first proposed a mathematical formulation for the design of trickling filters. Although the number of conventional trickling filter plants has been increasing each year, the percentage has been declining since 1957.

The development of plastic packings as high-rate trickling filter media has been one of the most significant advances in the field of biological wastewater treatment. The plastic medium was initially developed by the Imperial Chemical Industries, Ltd., in England during the early 1960s (27). Today on a worldwide basis, there is a large number full-scale biological filtration plants using plastic filter media and operating on wastes of widely differing character (15, 27, 28, 44–46).

Aerated-contact beds (also called submerged-contact aerators) were developed for more efficient continuous operation in the early 1900s (21). When a submerged-contact aerator is continuously operated, air is usually provided in sufficient volume to keep the waters

and slime surfaces aerobic and with sufficient intensity to tear away aging slime and solids accumulations for subsequent solid-water separation in a secondary settler.

Rotating disc systems have been used for biological wastewater treatment in Europe since 1960 (22). Research and development work has been conducted on this process in the United States since 1965 (22–24). In 1971 Torpey (25) received a patent on the invention of a new method and apparatus for substantially upgrading the operating efficiency of a biological rotating disc system. Torpey's invention uses forcibly rotating bodies that are partially submerged in the wastewater to provide surface area on which biological slimes develop for the purpose of removing pollutants from the wastewater.

The application of contact beds, trickling filters or rotating contact systems to wastewater treatment has been found profitable in areas where:

(a) Wastewater treatment personnel may be limited.
(b) Small flows exist.
(c) A plant effluent of from 20 to 30 mg/L of BOD_5 is acceptable.
(d) Land area requirements dictate height to be increased to achieve the designed bio-oxidation capacity.
(e) Intermittent discharges of toxic or inhibitory waste stream create shock conditions.
(f) Partial treatment or a specific treatment may be required on an industrial waste stream.

A new process, activated biofilters (ABF), which combines both attached and suspended-growth biological treatment systems, has been developed by Neptune Microfloc. The process entails the use of a redwood-medium trickling filter followed by aeration. The return sludge from the clarifier is recycled to the biocell along with the biocell underflow. Although the redwood medium is recommended by the manufacturer, plastic media may provide equally effective treatment (47).

Another development in the field of attached-growth wastewater treatment processes is the anaerobic filter (48). It is primarily a column filled with various types of filter media used for the anaerobic carbonaceous oxidation of organic matter in wastewater. The wastewater flows upward through the column, contacting the media on which anaerobic bacteria grow and are retained. Because the bacteria are retained on the media and not significantly washed off in the effluent, mean cell residence times on the order of 100 days can be obtained. Anaerobic filtering appears to be a viable process for the pretreatment of high-temperature and high-strength industrial wastes. It achieves high carbonaceous removal at relatively low operating cost and low sludge production compared to aerobic biological systems. Besides, it also produces methane as a useable end product according to Equation (4).

The emphasis of this chapter is on the principles of well-established aerobic attached-growth systems and the design of trickling filters. The design of rotating biological contactors, anaerobic filters and of activated biofilters is presented in detail elsewhere.

1.3. Microbiology and Ecology

A conventional microbial slime system is designed to contain an air-renewable surface to which a wastewater stream containing organic substrate and minerals is applied.

Microorganisms can metabolize the substrate in wastewater with net production of energy for growth or sufficient energy to maintain the existing population. Although the system is classed as an aerobic treatment device, it is not truly aerobic, but rather must be considered as facultative because the great majority of microorganisms in the system are facultative bacteria, such as *Pseudomonas*, *Alcaligenes*, *Flavobacterium*, *Micrococcus*, and so on, which live aerobically as long as dissolved oxygen is present and anaerobically when the oxygen is almost depleted. Aerobic bacteria, such as *Bacillus*, are found in the upper, aerobic slime surfaces (38). When an anoxic or anaerobic zone is formed in a thick slime layer, the obligate anaerobe, such as *Desulfovibrio* and some other sulfur-reducing bacteria have been isolated from the slime-medium interface (39). Under such circumstances the microbial slime system can develop odors and conceivably sloughing can occur as the result of gas generation in the interior slimes.

Fungi are aerobic microorganisms living in the aerobic zone of the slime and also decomposing the organic substrate in the wastewater. The contribution of fungi is significant only under low pH conditions or with unusual industrial effluents because the fungi cannot compete successfully with the bacteria for their food under normal environmental conditions.

Algae growing on the surface of a microbial slime system, such as trickling filter, are usually an inconsequential element of the microorganism's population, limited to illuminated surfaces. Algal microorganisms on a slime system are clearly tolerant of organic substances and high levels of carbon dioxide. Although algae add oxygen to the wastewater, they have been charged with responsibility for bed clogging and are considered to be troublesome from an operational standpoint (41).

The protozoa are the predominant small animals with all forms from the *Phytomastigophora* to *Suctoria*. The free-swimming ciliates predominate at the slime surface, while the stalked ciliates predominate in the lower regions. The primary function of protozoa is not to stabilize the waste, but to control the bacterial population in the system.

Higher animals including worms, snails and insect larvae feed on the lower forms of microorganisms in the microbial slime system and live in the upper aerobic areas. As a result, these higher animals can help to keep the bacterial population in a state of high growth or rapid food use. Filter flies are nuisance organisms in the trickling filters. Much of the early study of trickling filter populations was aimed at the control of these nuisance organisms by flooding, chlorination and the use of various pesticides.

A comprehensive description of the organisms found in trickling filters has been presented by Cooke (40), whose listing of various organisms can be found from ref. (49).

Trickling filters and similar microbial slime systems are short term retention devices that should not be expected to act as effective reduction devices for *S. typhosa, S. paratyphi*, and *Mycobacteria tuberculosis*. This is true also of pathogenic protozoa, such as *Entamoeba histolytica* (42).

When loaded lightly with carbonaceous substances, the trickling filter does some nitrification because of the presence of *Nitrosomonas* and *Nitrobacter*. However, when heavily loaded with carbonaceous matter, nitrification in a trickling filter or similar system may be absent or nominal.

2. THEORIES AND MECHANISMS

2.1. Transfer of Oxygen in Slime Layer and Liquid Film

The relationships of the supporting medium, the biological slime film, the waste liquid, and the atmosphere, as illustrated in Figure 9.1, have challenged the kinetic and mathematical skill of environmental engineers for many years. It was generally believed (8, 49) that the slime layer had an aerobic zone and an anaerobic zone, defined by the depth of penetration of oxygen in the slime layer. Further studies (18, 19) have reported that the biological slime film does not always consist of an aerobic layer and anaerobic layer. With dilute substrate, the respiration was found to be substrate-limited and high concentrations of oxygen were found throughout the slime mass. More concentrated substrates increased oxygen use; however, oxygen concentrations simply approached zero in the depth of the slime but were still above zero. A new term "anoxic" has been used by many environmental engineers to define an environment in which the dissolved oxygen concentration ranges from zero to about 1 mg/L. Accordingly it is believed that there would only be an aerobic zone in the slime layer when the substrate concentration is low, and there could be an aerobic zone and an anoxic zone in the slime layer when the substrate concentration in the liquid film is so high that oxygen concentrations are below 1 mg/L in the depth of the slime.

For a specific wastewater flow rate, the oxygen concentration gradient in a slime layer is a function of the dissolved oxygen concentration at the air-liquid interface and the substrate concentration. The latter affects the oxygen requirements of the slime layer. At low substrate concentrations in the liquid film, there will be a decrease in the oxygen requirements of the slime layer that should increase the oxygen concentration at the slime-liquid interface and reduce the mass flux of oxygen across the interface (43). The reduced oxygen requirement and increased oxygen concentration at the interface will then result in an increased depth of penetration of oxygen in the slime layer, and probably an aerobic zone throughout the full depth of the slime layer (18, 19). At high substrate concentration in the liquid film, the oxygen requirements of the slime layer approach a constant; then the mass flux of oxygen across the slime-liquid interface will be constant, and the thickness of the active portion (i.e., aerobic zone) of the slime layer for the entire bed depth will also be constant assuming this condition exists at any depth in the process bed (43).

The effect of hydraulic loadings on the transfer of oxygen in the slime layer in the liquid film has also been studied (36). As specified by Maier (50) and verified by Jank (43) a range for hydraulic loadings is normally encountered in full-scale operation of a trickling filter. For laminar flow, an increase in flow rate will result in an increase in liquid velocity and an increase in the mass flux of oxygen across the air-liquid and slime-liquid interfaces. The increased supply of oxygen at the slime-liquid interface will result in a greater depth of oxygen penetration in the slime layer, or a thicker active slime layer at a specific applied organic loading.

The depth of oxygen penetration is dependent on the molecular diffusion coefficient of oxygen in the slime layer, the rate of oxygen use, and the oxygen concentration at the slime-liquid interface. Jank and Dryman (36) have reported for a specific wastewater flow rate and substrate concentration, an oxygen concentration gradient will be established within

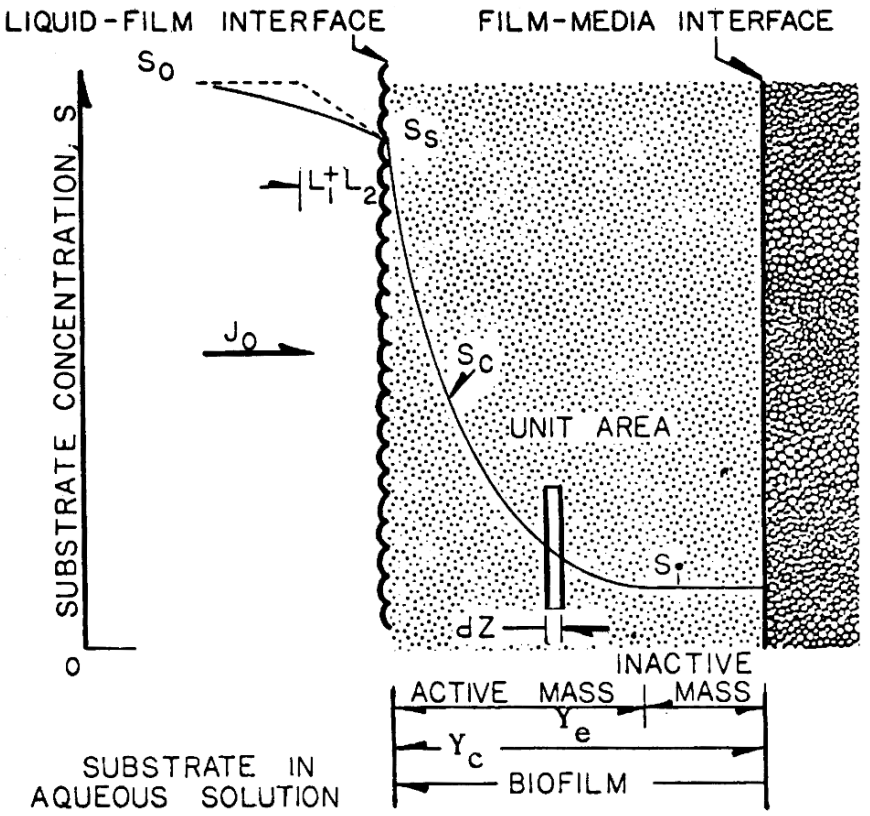


Fig. 9.4. Substrate profiles within a biofilm (*Source:* WPCF).

the slime layer that is directly related to the mass flux across the slime-liquid interface. Although the supply of oxygen at the slime-liquid interface can be either contributed by the dissolved oxygen content of the influent wastewater or transferred from the air to the liquid film as the wastewater flows across the slime layer, Jank (43) discovered that the quantity of oxygen contributed by the influent wastewater was negligible when compared to the oxygen requirement of the slime layer.

2.2. Transfer of Substrate in Liquid Film and Slime Layer

A conceptual illustration of the substrate concentration gradient within a slime layer (i.e., biofilm) as shown in Figure 9.4, has been proposed by Williamson and McCarty (51, 52). It is assumed that the rate of reaction is limited by a single substrate S. Let the substrate concentration outside the biofilm in the bulk liquid be S_0; at the biofilm surface, S_s; within the biofilm cellular matrix S_c and deep within the biofilm a constant limiting value, S_i. The substrate concentration gradient (dS_c/dz) at the slime-liquid interface where $z = 0$ has been shown by Whalen et al. (18) to be intermediate between low values (metabolism-limited case) and high values (diffusion-limited case). The biofilm depth to the point at which $S_c = S_i$ is termed the

effective depth, Y_e, and will contain those microorganisms actively metabolizing the substrate. The surface flux of the chemical species can be calculated from Fick's Law (53) as

$$J_0 = A_c D_w (S_0 - S_s)/Y \tag{8}$$

in which
A_c = biofilm area, cm^2
D_w = diffusion coefficient of the chemical species through water, cm^2/d
Y = the depth of a stagnant liquid layer outside the slime-liquid interface, cm
J_0 = surface flux of the chemical species, mg/d

Under steady-state conditions, the substrate concentration gradient will be retained in the biofilm and the mass flux of substrate across the slime-liquid interface will be equal to the total mass of substrate used by the active slime layer. If a biofilm is not metabolism-limited, then the substrate concentration within the depths of the biofilm will reach a minimum value of S_i, at which point bacterial metabolism stops. This situation occurs only in relatively thick biofilms. On the other hand, if the biofilm depth is restricted by either sloughing or hydraulic shear, then metabolism of the substrate may occur throughout the entire slime layer – this is a typical metabolism-limited case.

The mass transfer of substrate within the biofilm per unit area can also be described by Fick's Law as (51):

$$\frac{\partial S_m}{\partial t} = -A_c D_c (\partial S_c/\partial z) \tag{9}$$

in which
$\partial S_m/\partial t$ = rate of substrate mass transfer, mg/d
$\partial S_c/\partial z$ = substrate concentration gradient perpendicular to the surface plane, mg/cm^4
D_c = diffusion coefficient within the biofilm, cm^2/d

The rate of substrate use at any point within the biofilm is assumed to follow the Monod relationship (54, 55):

$$-(dS_c/dt) = kS_cX_c/(S_c + K_s) \tag{10}$$

in which
$-(dS_c/dt)$ = use rate of the rate-limiting substrate, mg/L
k = maximum use rate of the rate-limiting substrate, mg/d/mg
K_s = Monod half-velocity coefficient, mg/L
S_c = rate-limiting substrate concentration, mg/L
X_c = bacterial concentration within the biofilm, assumed to be constant with depth, mg/L

Through application of Equations (9) and (10) to the differential element of width dz shown in Figure 9.4, and combination of the mass transfer terms, a steady-state equation is derived

$$\frac{d^2 S_c}{dz^2} = \frac{k S_c X_c}{D_c (S_c + K_s)} \tag{11}$$

Equation (11) is a second-order nonlinear ordinary differential equation. Although it does not possess an explicit solution, it can be solved for the two limiting cases of the Monod equation (54). The biofilm surface flux and the biofilm substrate concentration for the limiting cases at $S_s \gg K_s$ and $S_s \ll K_s$ are presented elsewhere (51, 52). The biofilm model may be used to describe the use rate of any substrate by a biofilm if that substrate is both flux and substrate-limiting. Technical terms used in this chapter are all defined in the Nomenclature Section.

3. TYPES OF TRICKLING FILTERS

3.1. General Description

A trickling filter is a packed bed of media covered with slime over which wastewater is passed so that it "trickles" downward as a thin laminar film. Oxygen and organic substrates diffuse through the liquid film into the slime film where biological oxidation and synthesis occur. End products and sludges appear in the filter effluent.

3.2. Low-Rate, High-Rate, and Super-Rate Filters

Trickling filters are classified by hydraulic and organic loading as, low-rate, high-rate and super-rate (Figures 9.5 to 9.7). Super-rate filters are also referred to as roughing filters. There is no rigid dividing line between the various rate trickling filters (56). However, some rather general ranges are given in Table 9.1 to differentiate the various types of trickling filters by the hydraulic and organic loadings. Hydraulic loading is the rate of application of wastewater to the surface, usually expressed in millions of gallons per day per acre of surface area (mgad), gallons per day per square foot ($gal/d/ft^2$) or cubic meters per day per square meter ($m^3/d/m^2$). Organic loading is in pounds of BOD per acre-foot per day lb/ac.ft/d, pounds of BOD per cubic yard per day ($lb/yd^3/d$), pounds of BOD per 1,000 cubic foot per day ($lb/1000\,ft^3/d$) or kilograms of BOD per cubic meter per day ($kg/m^3/d$).

Accordingly, a low-rate filter is one designed for the applied loadings of not more than $0.4\,kg\ BOD/m^3/d$, ($25\,lb\ BOD/1000\,ft^3/d$) and $4.08\,m^3/m^2/d$ ($100\,gal/ft^2/d$). The low-rate filter, which is also referred to as standard rate filter, was the backbone of secondary biological treatment for over 50 years. The low-rate biological filter is about 1.5 to 2.5 m (5 to 8 ft) deep. Its rock media vary from 3.8 to 6.4 cm (1.5 to 2.5 in) nominal diameter and can be dosed with either fixed nozzle distributors or rotary distributors. Low-rate filters commonly have a dosing period of 3 minutes and a rest period of 6 minutes. The underdrains are usually sized to flow half-full at the design flow rate for adequate ventilation. Head loss through the filter may be 1.52–3.05 m (5 to 10 ft), which may be disagreeable if the plant site is too flat to permit gravity flow. Because the low-rate filter has a greater detention time and lower hydraulic and organic

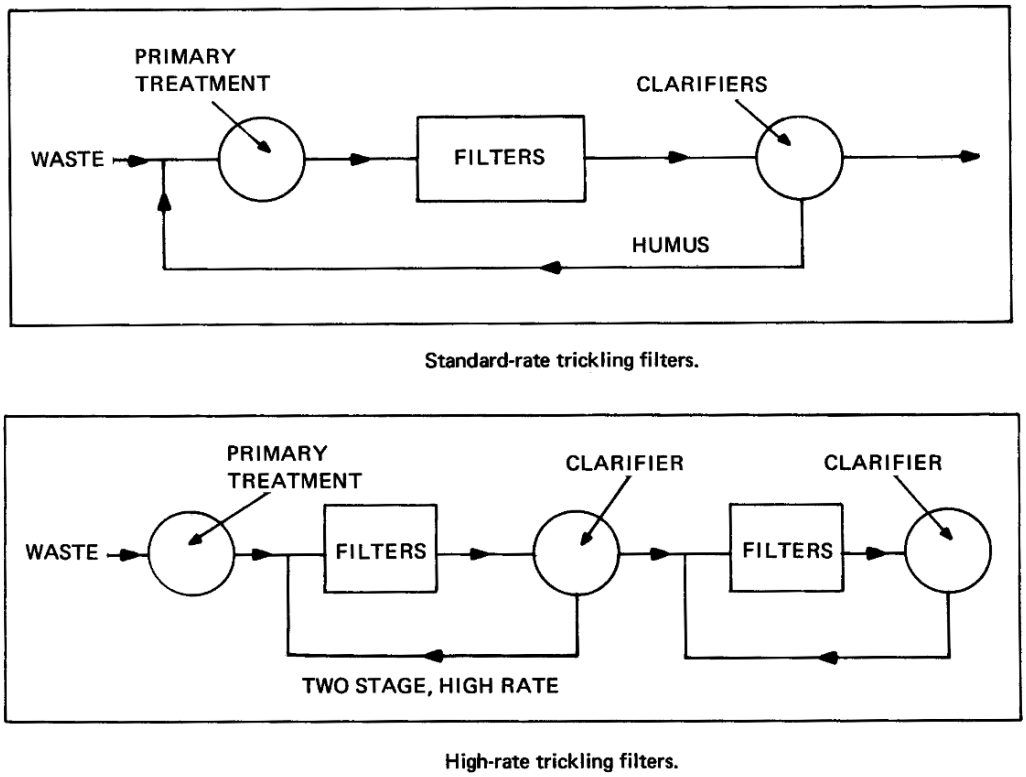


Fig. 9.5. Standard (low-rate) and high-rate trickling filters (*Source*: US EPA).

loadings, it produces a more highly nitrified effluent than the high-rate filter. The possible drawbacks of low-rate filters are:

(a) Odor problems caused by the septic sewage when weather is warm.
(b) The presence of filter flies.
(c) The occurrence of filter ponding.

For solving the odor problem, the low-rate filters should be located where the odors would not create a nuisance, or where the weather is not warm. When the filter fly persists as a nuisance, these nuisance organisms can be controlled by flooding, chlorination or the use of various pesticides (57, 58). Ponding of filters occurs when strong wastes are applied at low hydraulic loading rates. These difficulties are believed to be related to excessive amounts of bacterial growth that clog the trickling filters and can be controlled by higher flow rates to keep surfaces flushed. However, if both high organic and high hydraulic loadings are applied to a trickling filter, then the filter can no longer to be classified as a low-rate filter.

High-rate filters are normally designed for substantially higher organic and hydraulic loadings than low-rate filters. There are three principal types of high-rate filters depending

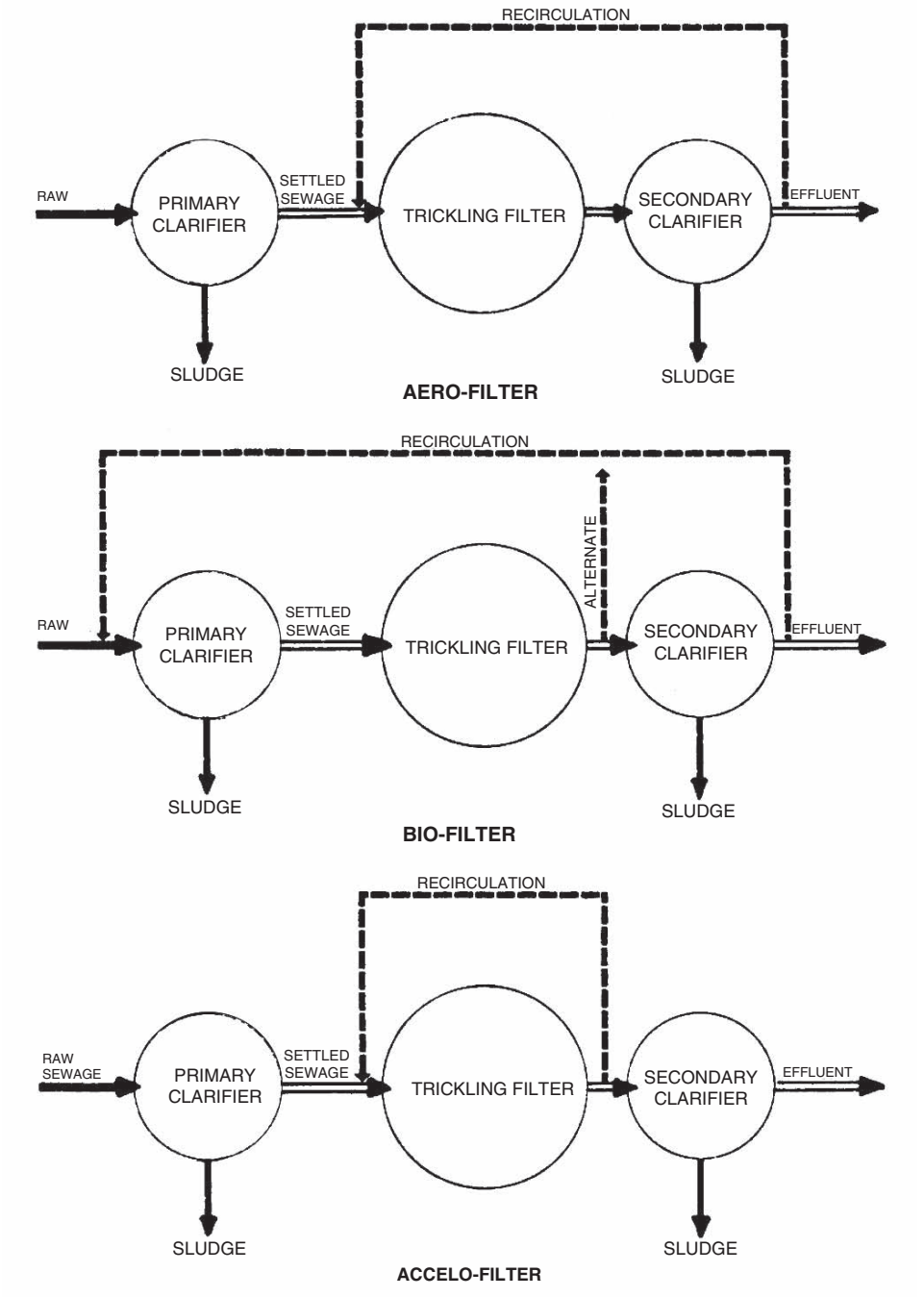


Fig. 9.6. Trickling filter variations (*Source*: US EPA).

Table 9.1
Design Comparison of Different Rate Filters

Design characteristics	Trickling filter classification				
	Standard (low) rate	Intermediate rate	High rate (rock media)	High-super rate (plastic media)	Super rate (for roughing)
Hydraulic loading					
gal/ft^2/d	25–100	100–230	230–1000	350–2100[a]	1400–4200
m^3/m^2/d	1.0–4.1	4.1–9.4	9.4–40.8	14.3–85.7	57.1–171.4
mgad	1.1–4.4	4.4–10	10.0–43.5	15.2–91.4	61.0–182.9
Organic loading					
lb BOD_5/1,000 ft^3/d	5–25	20–30	25–300	Up to 300	100 plus
kg BOD_5/m^3/d	0.08–0.40	0.32–0.48	0.40–4.80	Up to 4.80	1.60 plus
lb BOD_5/ac-ft/d	218–1089	871–1,307	1089–13,068	Up to 13,068	4356 plus
Recirculation (ratio)	Minimum (0)	Usually (0.5–3)	Always (0.5–3)	Usually	Not required
Filter flies	Many	Varies	Few	Few	Few
Sloughing	Intermittent	Varies	Continuous	Continuous	Continuous
Depth of bed, ft	5–8	5–8	4–8	Up to 40	3–20
BODs removal, %	80–85	50–70	65–80	65–85	40–65
Effluent nitrification	Well	Some	Nitrites	Limited	No

[a] Not including recirculation
Note: 1 gal/ft^2/d = 0.0408 m^3/m^2/d = 0.04354 mgad = 407.52 m^3/ha/d;
1 lb BOD_5/1000 ft^3/d = 0.016 kg BOD_5/m^3/d = 43.56 lb BOD_5/ac.-ft/d = 0.027 lb/BOD_5/yd^3/d.

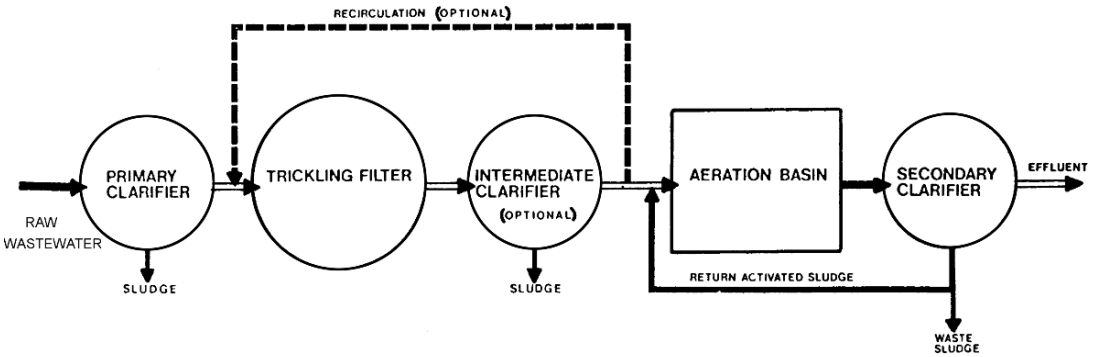


Fig. 9.7. Typical roughing filter installation (*Source*: US EPA).

on rate of feeding, recirculation, or loading. The three types are the Biofilter, the Accelo-Filter and the Aero-Filter, as indicated in Figure 9.6.

The biofilter is a relatively shallow filter, generally 1.2 to 1.5 m (4 to 5 ft) in depth, which uses recirculation of a portion of the filter effluent to the primary sedimentation basin for a second passage through the filter. Organic loadings of biofilters are in the order of 0.88 to

1.11 kg BOD/m^3/d (55 to 69 lb BOD/1000 ft^3/d) based on the strength of the primary tank effluent, with hydraulic loadings ranging from 9.38 to 28.15 m^3/m^2/d (230 to 690 gal/ft^2/d). If additional treatment is necessary to lower the BOD content in the effluent, a second-stage filter may be provided. By appropriate selection of flow patterns and recirculation ratios, it is possible to satisfy the desired degree of treatment.

The aero-filter, which has a relatively deep media bed of 1.83 to 2.74 m (6 to 9 ft), uses a low momentary rate of wastewater application to the filter by means of a special type of distributor designed for frequent "raindrop" applications over a maximum area of the filter at one time. Recommended organic loadings are from 1.11 to 1.19 kg BOD/m^3/d (69 to 74 lb BOD/1000 ft^3/d) and the hydraulic loading rate is more than 9.38 m^3/m^2/d (230 gal/ft^2/d) with recirculation if necessary to maintain this rate. It should be noted that recirculation is used only during periods of low wastewater flow, or only in amounts necessary to ensure proper operation of the distributor. BOD removal of single-stage treatment ranges from 63% to 78%. If additional treatment is desired, a second-stage filter may be provided, and with very strong organic wastewater intermediate clarification may be used. When two-stage treatment is used, the organic loading of first-stage filter is 1.65 to 1.80 kg BOD/m^3/d (2.75 to 3.0 lb BOD/yd^3/d) and the loading of second-stage filter is about 60% of the first-stage loading. Some of the arrangements used in the aero-filter plants are shown in Figure 9.6.

The accelo-filter, which is normally 1.83 to 2.44 m (6 to 8 ft) deep, uses direct recirculation of unsettled filter effluent back to the inlet of the distributor. The recommended organic loadings are in the same range as the biofilter, in the order of 0.88 to 1.11 kg BOD/m^3/d (55 to 69 lb BOD/1,000 ft^3/d) based on the strength of the primary clarifier effluent. Recirculation is used as in the biofilter to affect the desired degree of treatment. As with the biofilter, a large variety of flow patterns is possible, including use of a primary high-rate filter and a secondary low-rate filter.

Early investigators considered an applied hydraulic loading of 9.38 m^3/m^2/d (230 gal/ft^2/d) or more is necessary to flush organic solids from the filter media and prevent clogging. It was found later that serious clogging did not occur when dosing rates ranged from 4.08 to 9.38 m^3/m^2/d (100 to 230 gal/ft^2/d). Many trickling filters were then designed to operate in this so-called intermediate-rate range.

The super-rate filter is defined by its high hydraulic and organic loadings, as indicated in Table 9.1. The plastic media filters can be operated at the high/super-rate range for secondary treatment.

Either plastic media filters or rock media filters can be used as the so-called "roughing filters." A roughing filter (see Figure 9.7) is used to reduce the organic load in which subsequent treatment may be applied to the effluent or where intermediate treatment is required (59). Therefore, the roughing filler is generally installed advantageously ahead of activated sludge or any other secondary process. The filter design differs from other biological filters mainly because the determining factor is the high hydraulic loading as well as the high organic loading of certain wastewaters that are to be handled. Although roughing filters may give a high weight per unit volume of organic-load removal, their settled effluent will still contain substantial organics in terms of BOD_5.

Most of the primary filters with plastic packing in a two-stage system may be operated at high hydraulic and organic loadings because of the plastic media's light weight, requiring less support structure, less cost, and freedom from corrosion. Plastic super-rate filters generally have deep beds with plastic media that have large void spaces and are dosed continuously. The highest hydraulic loading found in the literature approached 469.2 $m^3/m^2/d$ (11,500 $gal/ft^2/d$) which, however, was applied to a shallow bed filter (49). The depth of the super-rate filter depends on the type of medium employed, however, the limitations in the medium available for use in super-rate filters were overcome by considerations outlined by Pearson (60). Recirculation is normally practiced to maintain efficiency and keep the slime film in a wetted condition.

3.3. Single- and Multi-Stage Trickling Filter Plants

A single-stage plant is one in which wastewater is passed through a single trickling filter; if there are two or more trickling filters, they would be operated in parallel. Wastewater may be, and usually is, recirculated through single-stage filters. Useful design charts were prepared by Baker and Graves (61) for single-stage filters using various design formulas. Although the single-stage filter is generally adequate and satisfactory for treating domestic wastewater, recent trends require an increase in detention time for handling occasional or constant overloaded conditions. The limitations of increasing depth are that the conventional rock filter media may not be able to support a very deep bed and a significant foundation structure must be provided. With the development of plastic media that are lighter in weight and possess proper ventilation characteristics, single-stage trickling filter plants using deep beds are being considered. Figures 9.5 and 9.6 show the flow diagrams of single-stage treatment.

In a two-stage filtration plant (i.e., double filtration plant), the effluent from a primary trickling filter, after the portion that is to be recirculated has been withdrawn for return to it directly or through the primary clarifier, passes through a secondary trickling filter. In other words, two-stage filtration means two biological trickling filters in series with or without intermediate clarifiers, followed by final clarification as shown in Figure 9.8. The development of the two-stage filtration plant grew out of necessity from overloaded conditions at treatment plants in the pre-World War II period. A typical example of overloaded condition would be an existing single-stage trickling filter with a relatively small volume and receives a strong waste high in BOD. In this case, the installation of a two-stage trickling filter plant could have been the solution to this problem. Many publications (8, 57, 61–66) present the formulas for designing two-stage trickling filter plants. The authors will present and summarize them in next section. As mentioned earlier, with the development of synthetic media, the use of a super-rate filter ahead of the existing trickling filter with rock media has been a popular practical solution.

Three-stage or tertiary treatment usually connotes activities concerned with nutrients control, primarily phosphorus and nitrogen. Although the tertiary filtration may be 80% to 100% superior to double filtration based on hydraulic advantages (67), it may be necessary only in exceptional cases. A factor in favor of the three-stage system is the development, under heterogeneous population of microorganisms, of selected strains of microorganisms in each filter stratum (49). Wastewater treatment plants having three-stage trickling filter processes always requires high capital investment, which is undesirable.

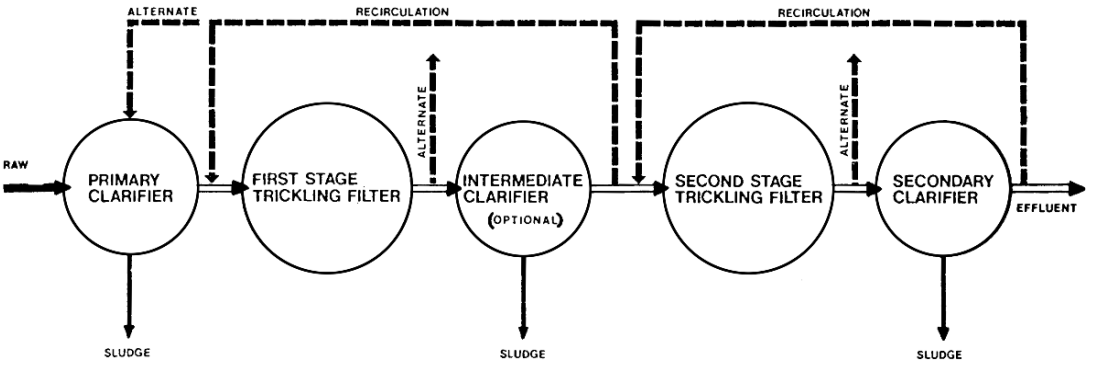


Fig. 9.8. Staging of filters (*Source*: US EPA).

4. PERFORMANCE MODELS AND DESIGN PROCEDURES

Although there are a variety of microbial slime systems, almost all performance models were developed for the design of conventional biological filters (26, 29–37, 68–78). Different design results, almost infinite in number, can be obtained for removing a given amount of biochemical oxygen demand (BOD) from a waste stream in accordance with a performance model when different values of bed depth, surface area, recirculation rate, hydraulic loading and wastewater temperature are assumed. The authors present the most common performance models in the following sections.

4.1. National Research Council Models

The National Research Council (NRC) compiled data for 34 operating rock trickling filters over an eight month period during World War II (26). The range of BOD removal in the field operations was between 75% and 95%. These field data are presented graphically in Figure 9.9. The average performance of a single-stage rock trickling filter based on these field data can be expressed by the empirical models proposed by the NRC in 1946.

$$E_1 = \frac{100}{1 + K_{\text{nrc}}(W/VF)^{0.5}} = (L_{0-}L_{\text{e}})/L_0 \tag{12}$$

$$F = \frac{1+R}{(1+0.1R)^2} \tag{13}$$

$$W = L_0 Q \tag{14}$$

$$R = Q_{\text{r}}/Q \tag{15}$$

in which

$K_{\text{nrc}} = 0.0085$

$E_1 = \%$ efficiency of BOD removal for process, including recirculation and sedimentation

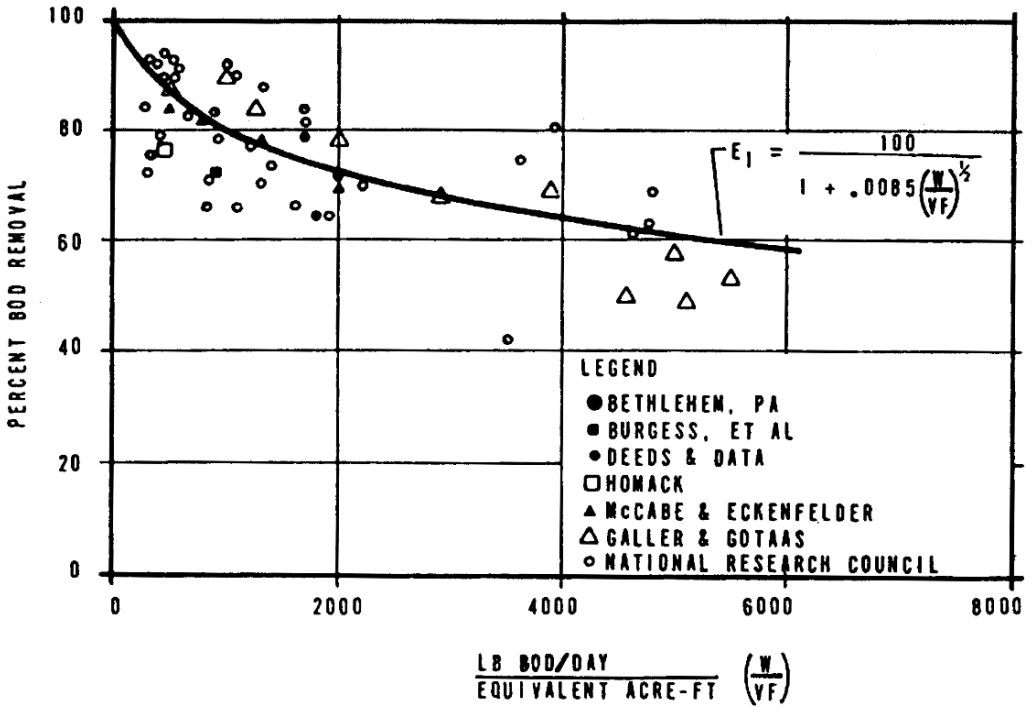


Fig. 9.9. Comparison of trickling filter operating data with NRC equation (*Source*: US EPA).

L_0 = BOD of influent, mass/volume, lb/MG
L_e = BOD of effluent, mass/volume, lb/MG
W = BOD loading to filter, mass/time, lb/d
V = volume of filter media, volume, ac.-ft
F = recirculation factor
R = recirculation ratio
Q = influent wastewater flow rate through the trickling filter, volume/time, MGD
Q_r = recirculation flow rate, volume/time, MGD

For the second-stage filter, the NRC model is:

$$E_2 = \frac{100}{1 + K_{nrc}(W'/VF)^{0.5}(1 - E_1)^{-1}} \tag{16}$$

$$W' = (1 - E_1)W \tag{17}$$

in which

E_2 = % efficiency of BOD removal for second-stage filtration, including recirculation and settling.
W' = BOD Loading to second-stage filter, mass/time, lb/d

4.2. Velz Model

In 1948 Velz (30) postulated that the BOD removal per unit depth of trickling filter was proportional to the BOD remaining.

$$dL/dD = -K_e L \tag{18}$$

which integrates to

$$L_D/L = e^{-K} e^{D} = 10^{-K_{10}D} \tag{19}$$

in which

D = the depth of the trickling filter, length, ft
L_D = the BOD remaining in the effluent at depth D, mass/volume, mg/L
L_0 = BOD of untreated wastewater, mass/volume, mg/L
L = applied BOD (mass/volume, mg/L) which is removable, not over $0.90\,L_0$
K_e = rate of BOD removal, base e
K_{10} = rate of BOD removal, base 10

When recirculation is used, the applied BOD_L may be determined from Equation (20):

$$L_a = \frac{QL_0 + Q_r L_e}{Q + Q_r} = \frac{L_0 + RL_e}{1 + R} \tag{20}$$

where

L_a = applied BOD (mass/volume, mg/L) after dilution by recirculation
L_e = effluent BOD, mg/L
R = recirculation ratio = Q_r/Q
Q_r = recirculation flow rate, volume/time, MGD
Q = influent wastewater flow rate through the trickling filter, volume/time, MGD

4.3. Upper Mississippi River – Great Lakes Board Model

The standards established by the Upper Mississippi River – Great Lake Board (33) in 1952 specify a maximum daily trickling filter loading of $1.77\,BOD_5/m^3$ ($110\,lb\,BOD_5/1000\,ft^3$) of filter, a filter depth of not less than 1.52 m (5 ft) nor more than 2.13 m (7 ft), a filter influent BOD concentration not to exceed three times the effluent BOD concentration and a hydraulic loading of not less than $9.88\,m^3/m^2/d$ (10 mgad) nor more than $29.65\,m^3/m^2/d$ (30 mgad). The standards were formulated into a performance model by Rankin in 1953 (68).

$$E = \frac{1 + \frac{Q_r}{Q}}{1.5 + \frac{Q_r}{Q}} \tag{21}$$

where E is the fractional efficiency of BOD removal.

4.4. Howland Models

In 1958 Howland (31) proposed that the rate of BOD removal was a function of contact time (t), giving the performance model

$$L_e/L_0 = e^{-k't} = \exp(-k't) \tag{22}$$

$$t = \frac{k''D}{Q^n} \tag{23}$$

in which n, k', and k'' are constants. Therefore, the remaining BOD in the effluent is obtained by substituting Equation (23) into Equation (22), yielding

$$L_e/L_o = \exp[-k_T(D/Q^n)] \tag{24}$$

in which k_T is the reaction rate at the wastewater temperature T, and n was determined to be 2/3. Besides, Howland (31) also introduced the effect of wastewater temperature on the reaction rate, k_T, in the BOD reduction equation

$$k_T = k_{20}\theta^{(T-20)} \tag{25}$$

in which T is the wastewater temperature, in degrees Celsius; k_{20} is the reaction rate at 20°C: and θ is the temperature coefficient equal to 1.035 according to Howland (31). The value of θ has been reported to vary from 1.020 to 1.072 by Eckenfelder (69).

4.5. Eckenfelder Models

Eckenfelder (32, 63, 64, 66) modified the Howland (31) and Schultze (70) models in 1961 to evaluate the effect of a decreasing amount of BOD removal per unit of depth with increasing trickling filter depth, resulting in a series of his performance models.

In a manner analogous to activated sludge under plug flow conditions, BOD removal can be related to the available biological slime surface and to the time of contact of wastewater with that surface.

$$L_e/L_o = \exp(-kX_v t) \tag{26}$$

in which

L_e = BOD remaining, mass/volume, mg/L
L_o = BOD in raw wastewater, mass/volume, mg/L
k = removal rate constant
X_v = volatile biological solids concentration, mass/volume
t = residence time, time

In a trickling filter, the mean residence time is defined as

$$T = CD^m/q^n \tag{27}$$

where

D = trickling filter depth, length, ft
q = hydraulic loading, volume/area/time, mgad

C, m, n = constants which are a function of the filter media and specific surface
$m = 1$ or 2 in most applications

The concentration of biological volatile solids, X_V, is a function of the specific surface of slime, a:

$$X_V = f(a) \tag{28}$$

where a = specific surface area (area of slime/volume of filter media). Therefore, the basic equation for BOD removal by a trickling filter with no recycle becomes:

$$L_e/L_o = \exp(-k' a D^m/q^n) \tag{29}$$

For a specific filter packing where, a, is known to be constant, Equation (29) becomes

$$L_e/L_o = \exp(-K D^m/q^n) \tag{30}$$

for a trickling filler with no recycle.

Equation (31) can be used for a trickling filter system with recycle.

$$L_e/L_a = \exp(-K D^m/q^n) \tag{31}$$

in which
L_a = BOD in raw wastewater following dilution with recycle flow, mg/L

$$K = k'a \tag{32}$$

When circulation is used, the influent BOD is diluted by recirculation flow. By a material balance, the BOD applied to the trickling filter (L_a) can be calculated by Equation (20). Because the trickling filter performance is a function of wastewater temperature, consideration must be given to temperature variation by adjustment of the reaction rate constant k or K according to Equation (25) when Eckenfelder's models are used for filter design.

4.6. Galler and Gotaas Model

In 1964 Galler and Gotaas formulated an empirical performance model for trickling filter design from a multiple regression analysis of data from pilot plants and existing trickling filter plants with effluent BOD, L_e, as the dependent variable (34, 72, 76)

$$L_e = \frac{0.46 L_a^{1.19}(1+R)^{0.28}(Q/A)^{0.13}}{(1+D)^{0.67} T^{0.15}} \tag{33}$$

in which
L_a = applied BOD (mass/volume, mg/L) after dilution by recirculation [see Equation (20)]
Q = influent wastewater flow rate through the trickling filter, volume/time, MGD
Q_r = recirculation flow rate, volume/time, MGD
A = trickling filter area, ac.
D = trickling filter depth, length, ft
T = wastewater temperature, °C

4.7. Biofilm Model

Figure 9.4 in Section 2.2 illustrates the biofilm model. Suspended organic wastes may be adsorbed onto the biofilm surface and hydrolyzed into smaller soluble substances that, together with other dissolved organics, diffuse through a relatively stagnant liquid layer into the biofilm. The oxygen required for biochemical oxidation of the organics must also diffuse into the biofilm at a rate proportional to the microorganisms' need. As oxygen and organics diffuse past the microorganisms in the biofilm, the microorganisms consume the organic wastes at a rate that is either a function of oxygen concentration or organic substrate concentration, depending upon which is limiting.

The flux of substrate into the biofilm (J, mg/cm^2/d) can be closely approximated by an equation of the form:

$$J = k_T S^p \tag{34}$$

where

S = substrate concentration (organics, ammonia or oxygen), mg/L
p = coefficient generally equal to 0.91 for oxygen, 0.94 for organic substrate and 0.97 for ammonia
k_T = rate coefficient (mg/cm^2/d) at wastewater temperature T°C
T = temperature, °C expressed by Equation (25), in which θ is equal to 1.039
k_{20} = 0.054 for organic substrate, 0.05 for ammonia, and 0.21 for oxygen, mg/cm^2/d

From a mass balance assuming plug flow through the reactor, Equation (34) can be integrated. The surface area in the biofilm (A_c, cm^2) is estimated by

$$A_c = aV \tag{35}$$

where a is the media surface per unit volume, cm^2/cm^3; and V is the volume of the attached growth media, cm^3. The integration of Equation (34) results in the following equation for substrate concentration in the reactor effluent:

$$S_e = [S_o^{(1-p)} - (1-p)k_T a V/Q]^{[1/(1-p)]} \tag{36}$$

where S_e is the effluent substrate concentration, mg/L; S_o is the bulk liquid substrate concentration, mg/L; p is a coefficient not equal to 1; and Q is the flow rate, cm^3/d.

4.8. US Army Design Formulas

In 1972, the US Army Corps of Engineers (US ACE) studied water problems in urban areas in addition to its traditional sanitary science role at recreation sites. Six years later, new design equations were developed by the US ACE for design of the plastic media trickling filters (74). Equations (37) to (40) are presented below for calculating the filter depth, filter surface area, media volume, and sludge production, respectively (15):

$$D = -\frac{q^n}{aK_{ace}} \ln \frac{L_e + L_e R}{L_o + L_e R} \tag{37}$$

where D = depth of filter, ft; q = hydraulic loading, gpm/ft^2; L_e = desired effluent BOD_5, mg/L; R = recirculation ratio = Q_r/Q; L_o = influent BOD_5, mg/L; a = specific surface area of the media, ft^2/ft^3; n = media factor, determined from laboratory; K_{ace} = reaction rate constant ranging from 0.0015 to 0.003, determined in the laboratory.

$$A = \frac{10^6 Q}{1440\, q} \tag{38}$$

where A = surface area of the filter, ft^2; and Q = average daily wastewater flow, MGD.

$$V = AD \tag{39}$$

where V = volume of filter media, ft^3

$$P_s = 8.34 Q L_o F_s \tag{40}$$

where P_s = sludge produced, lb/d; L_o = influent BOD_5, mg/L; and F_s = sludge production factor, lb solids/lb BOD_5. F_s value ranges from 0.42 to 0.65 lb solids/lb BOD_5.

4.9. US Environmental Protection Agency Model

A general US EPA model has been presented in a US EPA report (76) for designing all attached growth systems primarily to assess the removal phenomenon as a function of hydraulic loading rate per unit volume.

$$\frac{L_e}{L_o} = \exp -K_p \left(\frac{V}{695 Q}\right)^{0.5} \tag{41}$$

where

V = attached-growth media volume, ft^3
Q = wastewater design flow excluding recycle flow, MGD
L_e = reactor effluent BOD_5, mg/L
L_o = reactor influent BOD_5, mg/L
K_p = performance measurement parameter

$$K_p = 0.265 + \ln \frac{q_w}{20} \tag{42}$$

q_w = wastewater surface application rate (wetting rate), gpm/ft^2

The values of the performance measurement parameter (K_p) and the applicable wetting rates are presented below:

Filter Media	q_w, gpm/ft^2	K_p
Rock	0.1	0.15
Rock	0.2	0.18
Rock	0.3	0.20
Rock	0.4	0.22
Plastic	0.75	0.23

5. DESIGN AND CONSTRUCTION CONSIDERATIONS

A trickling filter has three principal components: (a) the distribution system that applies the wastewater to the filter media, (b) the filter medium that provides surface area for the microorganisms to grow, and (c) the underdrain system that supports the medium and provides drainage of the waste flow to a collection channel while permitting air circulation. The hydraulic considerations of three principal components, as well as related minor components are described in this section.

Wastewater may be distributed over the trickling filter by rotary distributors or other suitable devices permitting uniform distribution to the surface area. At design average wastewater flow, the deviation from a calculated uniformly distributed daily volume per unit area (such as gal/ft^2/d) shall not exceed $\pm 10\%$ at any point. Installations of motor-driven, rotary-type distributors have been used in filters ranging in size from 7.6 to 46 m (25 to 150 ft) in diameter.

There is a single conduit to convey the wastewater from primary settling tank to the distributor. Methods of conveying wastewater include gravity feed, dosing siphons, and pumping. When the filter is not designed for continuous dosing, the distributor is usually preceded by a pump or dosing tank and siphon.

All hydraulic factors involving proper distribution of wastewater on the filters should be carefully calculated. A minimum head of 0.6 m between low water level in siphon chamber and center of distributor arms is desirable. The head requirements of distributors are set by the manufacturers. The major head loss is the difference in elevation from the lowest water surface in the main underdrain channel. A minimum clearance of 0.15 m between distributor arms and filter media shall be provided. The head loss approximates 2.4 m (8 ft) for a filter 1.9 m (6 ft) deep and can be considerably greater for a deeper synthetic media filter.

It is important to know that bead losses chargeable to the trickling filter usually exceed the sum of all other head losses in the entire wastewater treatment plant. Compared with the activated sludge process, the trickling filter process requires much higher drop in static head, but requires less power.

Most trickling filter plants constructed in this country are circular and have rotary distributors. The majority of filters have reinforced concrete walls around the circumference, usually 0.2 to 0.3 m thick. The side walls are extended about 1 m to provide wind breaks.

The filter media may be crushed rock, slag, redwood, or specially manufactured synthetic plastic material. The media shall be durable, resistant to spalling or flaking, and be chemically and biologically inert. The characteristics of various filter media are shpwn in Table 9.2 (77).

Underdrains with semicircular inverts or equivalent should be provided and the underdrainage system shall cover the entire floor of the trickling filter. Inlet openings into the underdrains shall have an unsubmerged gross combined area equal to at least 15% of the surface area of the filter. The slope of underdrains shall be at least 1%. Effluent channels shall be designed to produce a minimum velocity of 0.6 m/s (2 ft/s) at average daily wastewater flow to the filter.

Ventilation of filter is important in maintaining the aerobic conditions necessary to secure effective biological treatment; therefore, adequate passageways at the bottom of the filters must be provided to permit free flow of air. Installation of vent stacks on the filter periphery,

Table 9.2
Summary of BOD removal characteristics of various media treating settled wastewater

Description	Specific surface ft^2/ft^3	Temperature range °C	Influent BOD range mg/L	Depth ft	Hydraulic loading range mgad	n[a]	K,[a] at 20°C
$1\frac{1}{2}$-in. flexirings	40.0	2–26	65–90	8	12.5–26.9	0.39	0.46
1-in. clinker	61.5	7–17	220–320	8	0.96–1.2	2.56	0.865
$2\frac{1}{2}$-in. clinker	37.4	7–17	220–320	6	0.96–1.2	0.84	0.685
1 -in. slag	60.0	7–17	220–320	6	0.96–1.2	0.30	0.865
$2\frac{1}{2}$-in. slag	33.0	7–17	220–320	6	0.96–1.2	0.75	0.640
1-in. rock	43.3	7–17	220–320	6	0.96–1.2	2.36	0.74
$2\frac{1}{2}$-in. rock	27.6	7–17	220–320	6	0.96–1.2	3.80	0.645
1-in. rounded gravel	44.5	7–17	200–320	6	0.96–1.2	3.00	0.625
$2\frac{1}{2}$-in. rounded gravel	19.7	7–17	220–320	6	0.96–1.2	5.40	0.57
Surfpac	28.0	24	200	21.6	31–350	0.50	0.395
Surfpac	28.0	24	200	12	62–250	0.45	0.33
$2\frac{1}{2}$- and 4-in. rock filter	15.0	24	200	12	31–94	0.49	0.275
$1\frac{1}{2}$- and $2\frac{1}{2}$ in. slag	42.0	7–17	112–196	6	5–12.5	1.0	0.87
1–3-in. granite	29.0	16–18	186–226	6	2–16	0.4	0.312
3/4-in. Raschig rings	75.8	16–18	186–226	6	2–16	0.7	0.55
1-in. Raschig rings	52.2	16–18	186–226	6	2–16	0.63	0.42
$1\frac{1}{2}$-in. Raschig rings	35.0	16–18	186–226	6	2–16	0.306	0.28
$2\frac{1}{4}$-in. Raschig rings	22.7	16–18	186–226	6	2–16	0.276	0.25
Straight block	28.2	16–18	186–226	6	2–16	0.345	0.2

[a] n and K are constants for Equation (31); m is equal to 1.
Source: Water and Wastes Engineering.

ventilating manholes, and discharge of filter effluent to the subsequent sedimentation basin in an open channel or partly filled pipes are methods employed to insure adequate ventilation. The Ten State Standards (33) require that inlet openings into the filter underdrains have an unsubmerged gross combined area equal to at least 15% of the surface area of the filter, and that the size of drains, channels, and pipe be such that no more than 50% of their cross-sectional area will be submerged under the design hydraulic loading. Synthetic media manufacturers often recommend 0.1 m^2 (1 ft^2) of ventilating area for each 3 to 4.6 m of plastic media filter's tower periphery for domestic wastewater. Consideration may also be given to a forced ventilation system.

6. PROCESS CONTROL CONSIDERATIONS

Trickling filters are usually preceded by preliminary and/or primary treatment facilities for preventing filter clogging, and followed by final sedimentation for efficient wastewater treatment.

For low-rate and some intermediate-rate trickling filters, the settled sludge in the final sedimentation unit is relatively stable, and periodic removal at 3 to 24 hours intervals depending upon operational conditions, is sufficient. Secondary sludge from a high-rate or super-rate trickling filter has a relatively highly oxygen demand, and therefore, it should be removed from the final sedimentation tank continuously.

Recirculation of filter effluent or final settler effluent is an important plant operation for equalization of wastewater strength as well as wastewater flow. Other advantages of recirculation include:

1. Increasing contact efficiency.
2. Seeding the filter throughout its depth with a large variety of microorganisms.
3. Improving operation of primary and secondary sedimentation during low wastewater flow periods by reducing septicity.
4. Improving distribution of wastewater over the filter surface.
5. Minimizing odors, ponding, and filter fly breeding by increasing hydraulic loading to encourage continuous sloughing and to reduce slime thickness.
6. Preventing biological growth from drying out during low wastewater flows.

Common engineering practice is to operate the high-rate trickling filter process for recirculation ratios of 0.5 to 4.0 considering ratios higher than 4.0 to be economically unjustifiable.

Various types of super-rate trickling fitters have different minimum wetting rates, that is, a rate of flow per unit area that will induce a biological slime throughout the depth of the media, The minimum wetting rate for the plastic media trickling filters typically ranges from 0.3 to $0.7\,\mathrm{L/m^2/s}$ (0.5 to $1.0\,\mathrm{gpm/ft^2}$) depending on the geometric configuration of the media.

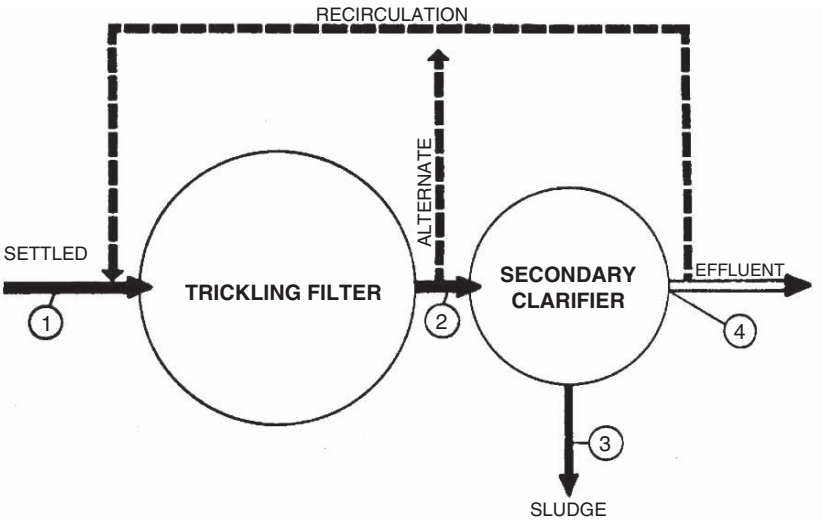
Effect of temperature on trickling filter performance can be expressed by the following relationship (55, 78)

$$E_{\mathrm{T}} = E_{20}\theta^{\mathrm{T}-20} \tag{43}$$

where θ is the temperature coefficient equals to 1.035; E_{T} = trickling filter treatment efficiency at temperature, T; E_{20} = trickling filter treatment efficiency at 20°C; and T = wastewater temperature, °C.

Extreme changes in pH can affect biological growth, thus reduce the filter's treatment efficiency. In general, pH value of wastewater should be controlled in the 6.5 to 8.5 range.

In addition to wastewater temperature and pH, other important process control parameters include settleable matter, suspended matter, dissolved oxygen (DO), biochemical oxygen demand (BOD), chemical oxygen demand (COD), flow, phosphorus, total solids, volatile solids, ammonia nitrogen, nitrite nitrogen and nitrate nitrogen. Figure 9.10 shows the minimum sampling and testing program for a typical trickling filter process (71). The settleable matter and suspended solids are two important tests for calculating the sludge removal and effectiveness of the sedimentation tank. The significance of the DO test in process control is in its determination of the availability of DO which is essential for biochemical oxidation of the organic waste. BOD and COD are mainly used for the determination of organic loading and biological treatment efficiency. A physical measurement of the in-plant flows is essential for computing hydraulic and organic loadings, detention periods, recycle flows, clarifier under



SAMPLE LOCATION FOR TYPICAL TRICKLING FILTER PROCESS

DESCRIPTION	SETTLED SEWAGE (PRIMARY EFFLUENT)	FILTER EFFLUENT	SLUDGE (UNDER FLOW)	SECONDARY CLARIFIER EFFLUENT	FREQUENCY 1 TO 5 MGD	FREQUENCY 5 MGD AND LARGER	TYPE OF SAMPLE	METHOD OF TEST	APPLICATION OF TEST
FLOW	CR	-	-	CR	-	-	-	-	P
BOD	X			X	1/W	2/W	C	AM	P
COD	X			X	1/W	2/W	C	AM	P
SUSPENDED SOLIDS TOTAL	X	X		X	1/W	2/W	C	AM	P
SUSPENDED SOLIDS VOLATILE	X			X	1/W	2/W	C	AM	P
SETTLEABLE SOLIDS	X	X		X	D	D	C	AM	P
AMMONIA	X			X	2/M	1/W	C	AM	P
NITRITE				X	2/M	1/W	C	AM	P
NITRATE				X	2/M	1/W	C	AM	P
PHOSPHORUS	X			X	2/M	1/W	C	AM	P
DO				X	1/W	2/W	G	AM	P
pH	X		X	X	D	D	G	AM	S
TEMPERATURE	X			X	D	D	G	AM	S
TOTAL & VOLATILE SOLIDS			X		1/W	2/W	C	AM	P
LOCATION OF SAMPLE	(1)	(2)	(3)	(4)					

CODE DESCRIPTION	
SAMPLE	S SURVEILLANCE
TEST RESULTS CALCULATED	W WEEK
(?) REMOTES SAMPLE LOCATION	M MONTH
C COMPOSITE	AM ANALYTICAL MEASUREMENT
D DALIY	CR CONTINUOUSLY RECORDED AND TOTALIZED
G GRAS	MM MAKE MEASUREMENT
P PROCESS CONTROL	PM PHYSICAL MEASUREMENT

Fig. 9.10. Sampling and testing program for trickling filter process (*Source*: US EPA).

flows, and so on. Reference to Figure 9.10 will indicate locations of typical in-plant flows that should be measured for process monitoring and control. Phosphorus is essential to biological growth: therefore, a phosphorus deficiency may lead to poor BOD removals. On the other hand, a high phosphorus concentration in the plant effluent may cause lake pollution, and thus must be reduced. Total and volatile solids tests are strictly used for in-plant process control. Ammonia, nitrite, and nitrate concentrations are valuable in evaluating the performance of the trickling filter process designed for biological nitrification and denitrification; however, they are not applicable to single-stage high-rate filters and super-rate roughing filters.

Several common design shortcomings and the ways to compensate for them are introduced below (15, 73):

(a) Fly nuisance caused by alternately wet and dry filter walls may be prevented by modifying ends of distributor arms to maintain continuously wet filter walls
(b) Odors resulting from poor ventilation of filter may be prevented by using forced air ventilation or covering the filter
(c) Ice buildup on filter media may be corrected by constructing a wind screen to protect the filter from prevailing wind, or covering the filter, pump sumps, and dosing tanks
(d) Clogging of distributor orifices caused by inadequate primary treatment may be prevented by improving grease and suspended solids removal in a primary sedimentation tank
(e) Trickling filter subject to clogging with leaves may be prevented by removal of nearby trees
(f) Excessive sloughing from trickling filter because of excessive organic loading may be prevented by using more trickling filters or expanding the trickling filtration plant to decrease the loading, and
(g) High flows through the final sedimentation tank carrying solids over the settler weir may be prevented by modifying the recirculation system so that trickling filter effluent (final settler influent) is recirculated directly

7. ENERGY CONSIDERATIONS

The trickling filter requires a hydraulic head for operation. Pumping energy requirements may be approximated by using the following equation:

$$E_{\mathrm{n}} = 1900QH \tag{44}$$

where E_{n} is the energy consumption in kwh/yr; Q is the total wastewater flow (including recirculation flow if any) in MGD; and H is the discharge head in ft. Equation (44) was derived based on the assumption that wire-to-water efficiency is equal to 60%.

For the typical head requirement of 10 ft for all rock media trickling filters, an energy requirement of 19,000 kwh/yr/MGD can be expected for low rate rock media filters assuming no recirculation flow and 95,000 kwh/yr/MGD can be expected for high rate rock media filters assuming an average recirculation ratio of 4/1. For the typical head requirement of 23 ft for plastic media trickling filters, and assuming an average recirculation ratio of 2:1, an energy requirement of 131,000 kwh/yr/MGD can be expected.

Additional energy requirements can be found elsewhere (75, 76, 79–81).

Table 9.3
Performance of various trickling filters treating domestic wastewater

Pollutant	Percent removal of various trickling filters		
	Low rate rock media[a]	High rate rock media[b]	High rate plastic media[b]
BOD_5	75–90	60–80	80–90
Suspended solids	75–90	60–80	80–90
Phosphorus	10–30	10–30	10–30
Ammonia nitrogen	20–40	20–30	20–30

[a]Single-stage configuration with primary and secondary clarification and no recirculation.
[b]For secondary treatment and using a single-stage configuration with filter effluent recirculation, and primary and secondary clarification.

8. APPLICATION, PERFORMANCE, AND RELIABILITY

Trickling filters are used for treating domestic and compatible industrial wastewaters amenable to aerobic bio-oxidation in conjunction with suitable pretreatment. Table 9.3 indicates the expected performance of single-stage trickling filters.

Rock media trickling filters have been used for nitrification; however, the treatment results obtained in field investigations are extremely variable. It is possible to produce nitrified effluent using plastic media trickling filters. In general, the design calculations for a single-stage nitrification filter have been based on an organic loading of 10 lb BOD_5/d/1000 ft^3 (0.16 kg BOD_5/m^3/d) of plastic media under peak month flow conditions. This value should result in an effluent ammonia-nitrogen concentration of about 3 mg/L, assuming the influent ammonia-nitrogen concentration equals about 20 mg/L.

The process can be expected to have a high degree of reliability if operating conditions minimize variability, and the installation is in a climate where wastewater temperatures do not fall below 13°C for prolonged periods. Mechanical reliability is high. The process is simple to operate.

As discussed earlier, the three-stage trickling filters can be used for carbonaceous oxidation, nitrification, and denitrification. If the third-stage trickling filter is intended to be used for denitrification, it must be covered. Additional carbon source, such as methanol, may be needed.

9. LIMITATIONS AND ENVIRONMENTAL IMPACT

An uncovered trickling filter is vulnerable to below freezing weather and its recirculation may be restricted during cold weather because of cooling effects. It is less effective in the treatment of wastewater containing high concentrations of soluble organics. It has only limited flexibility and control, and needs long recovery time with upsets. The process creates odor problems if improperly operated.

Settled sludge is withdrawn from the secondary clarifier at a rate of 2500 to 4000 L wet sludge/10^6 L of wastewater containing 180 to 320 kg (400 to 700 lb) dry solids.

10. RECENT DEVELOPMENT OF TRICKLING FILTERS

10.1. Treatment of Toxic and Volatile Organic Contaminants

Van der Hoek et al. (82) carried out bench-scale tests to remove polycyclic aromatic hydrocarbons, benzene, toluene, ethylbenzene, xylene and phenols from groundwater. The tests showed that trickling filters are less effective than upflow fixed film reactors in removing these compounds. Ring compounds are often toxic to biological action. A ring opening reaction of biorefractory pollutants to form organic acids of a biological nutrient was found to provide possible treatment of wastewater containing toxic compounds (83). Yang et al. (84) evaluated biodegradation efficacy of a dispersed diesel fuel under high salinity conditions using an aerobic, upflow submerged biofilter coupled with a trickling filter, which is used to capture and treat volatile organic compounds escaping from the biofilter caused by aeration. The total organic carbon (TOC) was removed with greater than 90% efficiency at a feed TOC concentration of 1,000 mg/L and a volumetric loading of 1.5 kg TOC/m^3/d.

Misra and Gupta (85) presented their initiation to explore the potential of a hybrid biological reactor, combining trickling filter and activated sludge process, to treat wastewater containing trichloroethylene (TCE) at ambient temperature at different hydraulic retention times (HRTs). The removal efficiency increased with decreasing flow rate with maximum TCE removal at an HRT of 28 hours (trickling filter 18 hours + activated sludge process 10 hours), achieving 81.5% efficiency excluding volatilization losses. Trickling filters have also been used to treat alkylbenzene sulphonate contaminants (86, 87). Langwaldt and Puhakka (88) reviewed on-site biological treatment for groundwater cleanup from industrial and agricultural chemicals. It is thought that a modified biological trickling filter was the better available pretreatment technology for BOD reduction from squid processing wastewater (89).

10.2. Metals and Biological Nitrogen Removal

Owing to increased demands on nitrogen removal, wastewater treatment using trickling filters is being extended with nitrifying reactors. Wik (90) carried out step and pulse response experiments on large pilot-scale nitrifying trickling filters to determine whether adsorption and desorption of ammonium as well as denitrification, can occur in trickling filters biofilm. He observed that adsorption and desorption caused a significantly slower transfer of the effluent ammonium concentration than expected from measured residence time distribution. He proposed a physically based model of simultaneous nitrification, denitrification, adsorption and desorption in the biofilm.

Schreff and Wilderer used an activated sludge process for COD removal as a first step and a trickling filter for subsequent nitrification to meet stricter requirements for nitrogen removal from municipal wastewater (91). Based on convective aeration caused by a fill and draw operational sequence, Lahav et al. (92) developed a vertical bed process for the removal of ammonium from secondary effluents that combines the advantages of the vertical wetlands concept with the high loading rates typically associated with trickling filters.

A maximal ammonium removal rate of 1100 g N/m^2 reactor/d was achieved using simulative effluents; however, an effective gravel size of 0.96 mm was clogged when using actual municipal secondary effluents. Two other media (2.46 mm and 4.31 mm) did not get clogged during the entire experimental period and a maximum removal load of 300 g N/m^2 reactor/d was achieved, much higher than typical rates reported for conventional vertical beds. For more information on nitrogen removal using trickling filter systems, the reader is referred to references (93–98).

Manganese removal, using a biological trickling filter, was found to be caused by both biological and chemical manganese oxidation. The continuous operational mode leads to higher percentage of manganese removal but lower throughput rates when compared with a sequencing batch reactor operation with the same feed concentration and retention time. When ammonia, iron and manganese are simultaneously present in the influent to a biological filter, there is no serious inhibition of manganese removal at low ammonia concentrations. At higher ammonia concentrations inhibition of manganese removal became substantial. The presence of iron affects both ammonia and manganese removal negatively, while ammonia and manganese do not significantly affect iron removal (99). Rapid oxidation and accretion of iron onto the large surface area media posses as a potential passive treatment option for mine wastewaters. Two pilot-scale reactors containing different large surface area plastic media trickling filters showed that the oxidation and accretion may be a promising alternative passive technology (100).

10.3. Structure of Biofilms and Characterization of Filter

Wastewater treatment performance is still typically defined in terms of nonspecific parameters such as biochemical oxygen demand and total suspended solids. Marquet et al. (101) employed high performance size exclusion chromatography to study the evolution of the dissolved pollutant fraction of wastewater through treatment and characterize trickling filter effluent by particle size distribution. It was shown that the method could be a valuable tool to assess the performance of wastewater treatment processes. Particle size distribution in effluent of trickling filters has also been characterized by Schubert and Gunthert (102).

In the trickling filter process, knowledge of the hydrodynamic behavior is essential for accurate mechanistic modeling. Numerous measurements including the mean residence time, the free draining volume and the residence time distribution have been performed; however, there is still a lack of data for the full-scale process. Seguret et al. (103) studied eight full scale trickling filters receiving urban wastewater to establish correlation for stone-packed filters and for vertical and random plastic-packed filters.

Morgenroth and Wilderer (104) found different detachment patterns have a significant influence on organism distribution within the biofilm and on overall process performance. Therefore, they questioned whether results from laboratory experiments in laminar flow channels (with constant erosion) can be used to describe structure and function of biofilms in full-scale biofilm reactors with large time dependent fluctuations of the biofilm thickness.

It is necessary to understand the mechanism of macromolecule degradation by bacteria in trickling filter effluent. The majority of protein hydrolytic activity in trickling filter effluent (>80%) was associated with suspended cells, although it is likely that substantially higher hydrolytic activity would be produced by the biofilm itself than was produced by the sloughed and recycled cells present in trickling filter effluent (105).

In PETRO (pond enhanced treatment and operation) system (106), microalgae in particular play an important role within the trickling filter biofilm consortium. Their stress-induced biofilm slime production under heterotrophic conditions in the trickling filter appears to greatly facilitate flocculation of solids which results in high performance of the trickling filter, the clarifier and the system as a whole.

During recent years the structure of biofilms from many different environments has been documented and evaluated by use of a broad variety of microscopic, physicochemical and molecular biological techniques, revealing a generally complex 3D structure (107). The expanding field of molecular techniques not only allows more and more detailed documentation of the spatial distribution of species, but also of functional activities of single cells in their biofilm environment. These new methods will certainly reveal new insights in the mechanisms involved in the developmental processes involved in the formation and behavior of biofilms.full-scale nitrifying trickling filter treating municipal wastewater has been investigated with microbiological methods using fluorescence in situ hybridization with oligonucleotide probes in combination with confocal laser scanning microscopy and mathematical modeling using a dynamic multi-species biofilm reactor model (108). They concluded that biofilm is depth dependent, a trend of a decrease with filter depth of the amount of biofilm, the proportion of all bacteria and the total amount of ammonia oxidizing bacteria. Pedersen and Arvin (109) used a specific oligonucleotide 16S ribosomal RNA as probe to target the toluene degradation.

Mathematical models are useful tools in the prediction of the system response to operational changes (110–115). Several models have been proposed to simulate nitrification in full-scale trickling filters (111, 112). Another approach used to characterize trickling filter performance is nuclear magnetic resonance (NMR) (113), which is used to better understand the distribution of flow velocities within the packed column.

10.4. Upgrading and Retrofitting

Trickling filters are a critical part of the nutrient removal processes and contribute to their flexibility and stability. Retrofitting an existing trickling filter plant to BNR has been recently undergone in Australia and has proved to be cost effective (116). A compact process based on tertiary nitrification in trickling filters and recirculation of trickling filter effluent made it possible to double the secondary settler capacity in existing activated sludge plant. It indicated that denitrification could be achieved without any extension of the existing activated sludge plant (117). The largest treatment plant in Vancouver, Canada was upgraded and commissioned at a cost of 470 million Canadian Dollars in 1998 (118). New large trickling filters, aerated solids contact tanks, secondary clarifiers and an array of other related facilities and improvements were added to the plant.

11. DESIGN EXAMPLES

11.1. Example 1.

A plastic media trickling filter serving a flow of 1.0 MGD is to be designed with the following given data:

(a) Hydraulic loading to filter (q) = 1.0 gpm/ft^2
(b) Influent BOD_5 concentration (L_0) = 200 mg/L
(c) Desired effluent BOD_5 concentration (L_e) = 15 mg/L
(d) Recirculation ratio (R) = 1.0
(e) Specific surface area (a) = 30 ft^2/ft^3
(f) Reaction rate constant (K_{ace}) = 0.0022 ft/min
(g) Sludge production factor (F_s) = 0.45 lb solids/lb BOD_5
(h) Media factor (n) = 0.5

Determine the following design parameters using the US Army design formulas:

(a) Filter depth (D), ft (Note: The filter depth must be checked against $D \leq 30$ ft. If $D > 30$ ft, select a lower hydraulic loading, q, and recalculate D)
(b) Surface area of filter (A), ft^2
(c) Volume of filter media (V), ft^3
(d) Sludge produced (P_s), lb/d

Solution

(a) Calculate the filter depth with Equation (37):

$$D = -\frac{q^n}{aK_{ace}} \ln \frac{L_e + L_e R}{L_o + L_e R} \tag{37}$$
$$= -(1.0^{0.5})/(30 \times 0.0022) \ln[(15 + 15 \times 1)/(200 + 15 \times 1)]$$
$$= 29.84 \text{ ft, say } 30 \text{ ft.}$$

(b) Calculate the surface area of filter

$$A = \frac{10^6 Q}{1440 q} \tag{38}$$
$$= 10^6(1)/(1440(1)$$
$$= 694.44 \text{ ft}^2\text{, say } 695 \text{ ft}^2$$

(c) Calculate the volume of filter media:

$$V = AD \tag{39}$$
$$= (695 \text{ ft}^2)(30 \text{ ft})$$
$$= 20{,}850 \text{ ft}^3$$

(d) Calculate the sludge produced:

$$\begin{aligned} P_s &= 8.34 Q L_0 F_s \\ &= 8.34(1.0)(200)(0.45) \\ &= 750\,\text{lb/d} \end{aligned} \tag{40}$$

11.2. Example 2

Two rock media trickling filters are to be designed for carbonaceous organic removal using the General US EPA Model, and assuming the following given data:

(a) Average daily wastewater flow (Q_{ave}) = 1 MGD
(b) Wastewater temperature (T) = 10°C
(c) Design flow (Q) = peak monthly flow (Q_{pm}) = $1.45 Q_{ave}$
(d) Peak daily flow (Q_{pd}) = $3.5 Q_{ave}$
(e) Desired effluent BOD_5 (L_e) = 20 mg/L under maximum month conditions
(f) Influent to the trickling filter = primary effluent
(g) Surface application rate (q_w) = 0.16 gpm/ft^2
(h) Temperature coefficient (θ) = 1.02
(i) Primary effluent BOD_5 = 135 mg/L
(j) Recirculation ratio of 1:1 for peak month conditions (Q_r/Q)
(k) Number of filters = 2
(l) Depth of filter (D) = 6 ft

Determine the following design parameters:

(a) Recirculation ratio (Q_r/Q_{ave}) when $Q_r = Q_{pd} - Q_{ave}$
(b) Recirculation pump requirements
(c) Total filter media volume, ft^3
(d) Total filter surface area, ft^2
(e) Diameter of each filter, ft
(f) Average and peak month BOD_5 loadings, lb/1,000 ft^3/d
(g) Hydraulic loading including recirculation for peak month conditions, gpm/ft^2

Solution

(a) Average daily flow (Q_{ave}) = 1 MGD
Peak daily flow (Q_{pd}) = $3.5 Q_{ave}$ = 3.5 MGD
$Q_r = Q_{pd} - Q_{ave} = 3.5 - 1 = 2.5$ MGD
Recirculation ratio (Q_r/Q_{ave}) = 2.5:1
Recirculation ratio (Q_r/Q_{pm}) = 1:1 (given)

(b) Recirculation pumps should be selected based on the highest pumping capacity (now, 2.5 MGD total), and one unit is needed for standby.

Select 3 recirculation pumps at 2 MGD each. Two are for the two filters, and the third one is the standby.

(c) Calculate the performance measurement parameter:

$$K_p = 0.265 + \ln(q_w)/20 \tag{42}$$
$$= 0.265 + \ln(0.16)/20$$
$$= 0.173 \text{ at } 20^\circ\text{C}$$
$$K_p \text{ at } 10^\circ\text{C} = (K_p \text{ at } 20^\circ\text{C})\theta^{(10-20)} \tag{25}$$
$$= 0.173 \times 1.02^{-10}$$
$$= 0.14$$

Calculate V with Equation (41):

$$L_e/L_o = \exp[-K_p(V/695Q)^{0.5}]$$
$$20/135 = \exp[-0.14(V/695 \times 1.45)^{0.5}]$$
$$V = 188{,}000\,\text{ft}^3$$

(d) Total surface area $A = V/D = 188{,}000/6 = 31{,}300\,\text{ft}^2$

(e) Two trickling filters.

$$\text{Area of each filter} = 31{,}300/2 = 15{,}650\,\text{ft}^2$$
$$15{,}650 = (d/2)^2 \times 3.14$$
$$\text{Diameter}(d) = [(15{,}650/3.14)4]^{0.5} = 141.2\,\text{ft}$$

(f) Under peak monthly flow conditions:

$$\text{BOD}_5 \text{ loading} = 1.45(135)(8.34)/188 = 8.7\,\text{lb}/1{,}000\,\text{ft}^3/\text{d}$$

Under average daily flow conditions:

$$\text{BOD}_5 \text{ loading} = 1.0(135)(8.34)/188 = 6.0\,\text{lb}/1{,}000\,\text{ft}^3/\text{d}$$

(g) For a recirculation ratio of 1:1 for peak month conditions, the hydraulic loading including recirculation flow is

$$q = 695(1.45 + 1.45)/31{,}300 = 0.065\,\text{gpm}/\text{ft}^2$$

11.3. Example 3

Two plastic media trickling filters are to be designed for carbonaceous organic removal using the General US EPA Model, and assuming the following given data:

(a) Average daily wastewater flow (Q_{ave}) = 1 MGD
(b) Wastewater temperature (T) = 10°C
(c) Design flow (Q) = peak monthly flow (Q_{pm}) = $1.45Q_{ave}$
(d) Peak daily flow (Q_{pd}) = $3.5Q_{ave}$
(e) Desired effluent BOD_5 = 22 mg/L under peak month conditions
(f) Influent to the trickling filters = primary effluent
(g) Primary effluent BOD_5 concentration = 135 mg/L
(h) Performance measurement parameter at 20°C(K_p) = 0.2

(i) Temperature coefficient (θ) = 1.02
(j) Number of filters = 2
(k) Depth of filter (D) = 21 ft
(I) Recirculation ratio (Q_r/Q) = 2.5 : 1 for peak month conditions

Determine the following design parameters:

(a) Recirculation ratio (Q_r/Q_{ave}) assuming the hydraulic loading including recycle is kept constant
(b) Recirculation ratio (Q_r/Q_{pd}) assuming the hydraulic loading including recycle is kept constant
(c) Recirculation pump requirements
(d) Total filter media volume, ft^3
(e) Total filter surface area, ft^2
(f) Average and peak month BOD_5 loading, $lb/1{,}000\,ft^3/d$
(g) Total hydraulic loading, gpm/ft^2

Solution

(a) Average daily flow (Q_{ave}) = 1 MGD
Peak monthly flow (Q_{pm}) = $1.45Q_{ave}$ = 1.45 MGD
Peak daily flow (Q_{pd}) = $3.5Q_{ave}$ = 3.5 MGD
Given: $R = Q_r/Q_{pm} = 2.5$

$$Q_r + Q_{pm} = \text{constant} = (2.5 + 1)Q_{pm} = 5.08\,\text{MGD}$$

Recirculation ratio (Q_r/Q_{ave}) = (5.08 − 1) = 4 : 1
(b) Recirculation ratio (Q_r/Q_{pd}) = (5.08 − 3.5)/3.5 = 0.45 : 1
(c) Two recirculation pumps at 2.5 MGD each will satisfy the total peak pumping capacity of 4 MGD.
One additional pump is needed as a standby.
(d) Given: K_p at 20°C = 0.2, and $\theta = 1.02$

$$K_p \text{ at } 10°C = 0.2 \times 1.02^{(10-20)} = 0.16$$

$$L_e/L_o = \exp[-K_p(V/695Q)^{0.5}]$$

$$22/135 = \exp[-0.16(V/695 \times 1.45)^{0.5}]$$

$$V = 130{,}000\,ft^3$$

(e) Given: D = 21 ft

$$\text{Total surface area}(A) = V/D = 130{,}000/21 = 6{,}200\,ft^2$$

(f) Under peak monthly flow conditions:

$$BOD_5 \text{ loading} = 1.45(135)(8.34)/130 = 12.5\,lb/1{,}000\,ft^3/d$$

Under average daily flow conditions:

$$BOD_5 \text{ loading} = 1(135)(8.34)/130 = 8.7\,lb/1{,}000\,ft^3/d$$

(g) Hydraulic loading for $R = 2.5$ under Q_{pm} conditions:

$$q = 695(1.45 + 2.5 \times 1.45)/6200 = 0.57\,gpm/ft^2$$

(Note: hydraulic loading is assumed to be constant regardless of conditions)

11.4. Example 4

Four plastic media trickling filters are to be designed for both carbonaceous oxidation and nitrification. Given data and design criteria are presented below:

(a) Average daily wastewater flow (Q_{ave}) = 10 MGD
(b) Wastewater temperature (T) = 10°C
(c) Design flow (Q) = peak monthly flow (Q_{pm}) = $1.38\,Q_{ave}$
(d) Peak daily flow (Q_{pd}) = $2.5\,Q_{ave}$
(e) Influent to the trickling filters = primary effluent
(f) Primary effluent BOD_5 concentration = 135 mg/L
(g) Primary effluent ammonia-nitrogen concentration = 20 mg/L
(h) Performance measurement parameter at 20°C(K_p) = 0.2
(i) Temperature coefficient (θ) = 1.02
(j) Single-stage nitrification system with both primary and secondary clarifiers
(k) Number of filters = 4
(l) Depth of filter (D) = 21 ft
(m) Recirculation ratio (Q_r/Q) = 4.8 : 1 for peak month conditions
(n) Use general US EPA models

Determine the following design parameters:

(a) Recirculation ratio (Q_r/Q_{ave}) assuming the hydraulic loading including recycle is kept constant
(b) Recirculation ratio (Q_r/Q_{pd}) assuming the hydraulic loading including recycle is kept constant
(c) Recirculation pump requirements
(d) Total filter media volume, ft^3
(e) Total filter surface area, ft^2
(f) Average and peak month BOD_5 loadings, $lb/1{,}000\,ft^3/d$
(g) Total hydraulic loading, gpm/ft^2
(h) Filter diameter, ft
(i) Effluent BOD_5 concentrations at 20 and 10°C

Solution

(a) Peak monthly flow (Q_{pm}) = $1.38\,Q_{ave}$ = 13.8 MGD
Peak daily flow (Q_{pd}) = $2.5\,Q_{ave}$ = 25 MGD
Given: $R = Q_r/Q_{pm} = 4.8$
$Q_r + Q_{pm}$ = constant = $(4.8 + 1)\,Q_{pm}$ = 80 MGD
Recirculation ratio (Q_r/Q_{ave}) = $(80 - 10)/10 = 7:1$
(b) Recirculation ratio (Q_r/Q_{pd}) = $(80 - 25)/25 = 2.2:1$
(c) The highest total pumping capacity is 70 MGD and there are four filters. Therefore, four recirculation pumps at 20 MGD each are required. One additional pump at 20 MGD is needed as a standby.
(d) The design calculations for a single-stage nitrification filter are based on an organic loading of 10 lb $BOD_5/d/1000\,ft^3$ of plastic media under peak month flow conditions.

$$BOD_5 \text{ load} = (135 \times 8.34)(13.8) = 15{,}537.42\,lb/d$$

$$\text{Total filter media}(V) = 15537.42/(10/1000) = 1{,}553{,}700\,ft^3$$

(e) Total filter surface area

$$A = V/D = 1{,}553{,}700/21 = 74{,}000\,\text{ft}^2$$

(f) Under peak monthly flow conditions:

$$\text{BOD}_5 \text{ loading} = 10\,\text{lb}/1000\,\text{ft}^3/\text{d(Given)}$$

Under average daily flow conditions:

$$\text{BOD}_5 \text{ loading} = (135 \times 8.34)(10)/1553.7 = 7.25\,\text{lb}/1000\,\text{ft}^3/\text{d}$$

(g) Total hydraulic loading

$$= (695\,\text{gpm/MGD})(80\,\text{MGD})/(74{,}000\,\text{ft}^2) = 0.75\,\text{gpm/ft}^2$$

(h) Area for each filter

$$\begin{aligned} A &= 74{,}000/4 = 18{,}500\,\text{ft}^2 \\ d &= (A/3.14)4]^{0.5} \\ &= [(18{,}500/3.14)4]^{0.5} \\ &= 153.5\,\text{ft} \end{aligned}$$

(i) Effluent BOD_5 at 20°C (L_e):

$$\begin{aligned} L_e &= L_o \exp[-K_p(V/695Q)^{0.5}] \\ &= 135\ \exp[-0.20(1{,}553{,}700/695 \times 13.8)^{0.5}] \\ &= 10.6\,\text{mg/L} \end{aligned} \tag{41}$$

Effluent BOD_5 at 10°C(L_e):

$$\begin{aligned} L_e &= 135\exp[-0.20(1.02)^{10-20}(1{,}553{,}700/695 \times 13.8)^{0.5}] \\ &= 16.7\,\text{mg/L} \end{aligned}$$

11.5. Example 5

Derive a working equation from original Eckenfelder Model for designing a trickling filter plant with recirculation around the filter.

Solution

The original Eckenfelder Model for a trickling filter with recirculation is

$$L_e/L_a = \exp(-KD^m/q^n) \tag{31}$$

The BOD_5 applied to the filter can be expressed by

$$L_a = (L_o + RL_e)/(1 + R) \tag{20}$$

The soluble BOD removal relationship for filters with recirculation can be derived by substituting Equation (20) into Equation (31):

$$L_e/[(L_o + RL_e)/(1 + \mathrm{R})] = \exp(-K D^m/q^n)$$

$$L_e = (L_o + RL_e)[\exp(-KD^m/q^n)]/(1 + R)$$

$$L_e/L_o = [1 + R(L_e/L_o)][\exp(-KD^m/q^n)]/(1 + R)$$

Finally the working design equation for filters with recirculation is:

$$\frac{L_e}{L_o} = \frac{\exp(-KD^m/q^n)}{(1 + R) - R[\exp(-KD^m/q^n)]}$$

where L_e = effluent BOD_5 concentration, mg/L; L_o = trickling filter influent BOD_5 concentration before dilution with recycle flow, mg/L; K = removal rate constant; $R = Q_r/Q$ = recirculation ratio; D = filter depth, ft; $m = a$ constant between 1 and 2 in most applications; $n = a$ constant to be determined in the laboratory; and q = hydraulic loading excluding recirculation, mgad.

11.6. Example 6

A trickling filter experimental plant can be installed to determine the constants of a specific filter medium in Eckenfleder's Model. The investigation is necessary to provide information in the development of new design criteria when a new filter medium is to be used. The trickling filter model built by Balakrishnan et al. (77) was a 20-in. diameter, 9-ft deep filter with an air sparger to provide a uniform air distribution from the bottom and distribution plates on top for uniform hydraulic loading. The model filter was packed to a depth of 8 ft using 1.5-in. polypropylene Flexiring media, manufactured by Koch Engineering, New York. The filter medium had 96% of free space and 40 ft^2/ft^3 of specific surface. The model filter was acclimated with the settled domestic wastewater and operated at 14°C as a secondary treatment system without recirculation. Suppose the same model filter was used. Samples were collected at various depths in the filter for laboratory analysis and the samples were settled for 30 min and filtered through a Whatman No. 42 filter paper before testing of effluent BOD_5. The average percent BOD_5 remaining was then calculated by:

$$\text{Average } BOD_5 \text{ remaining}(\%) = (L_e/L_0)100$$

Figure 9.11 shows the experimental data obtained at three different filtration rates: 0.2, 0.3, and 0.43 gpm/ft^2. Determine the constants K, m and n in Equations (30) and (45).

Solution

Equation (30) can be rewritten as:

$$\ln(100L_e/L_0) = \ln 100 - Kq^{-n}D^m \tag{46}$$

In Figure 9.11, the slope is $-Kq^{-n}$, the ordinate is $\ln(100L_e/L_o)$, and the abscissa is simply D, not D^m.Because the experimental data can be expressed as three straight lines on the semilogarithmic plot, one can reasonably assume that the value of m equals to 1. The equation

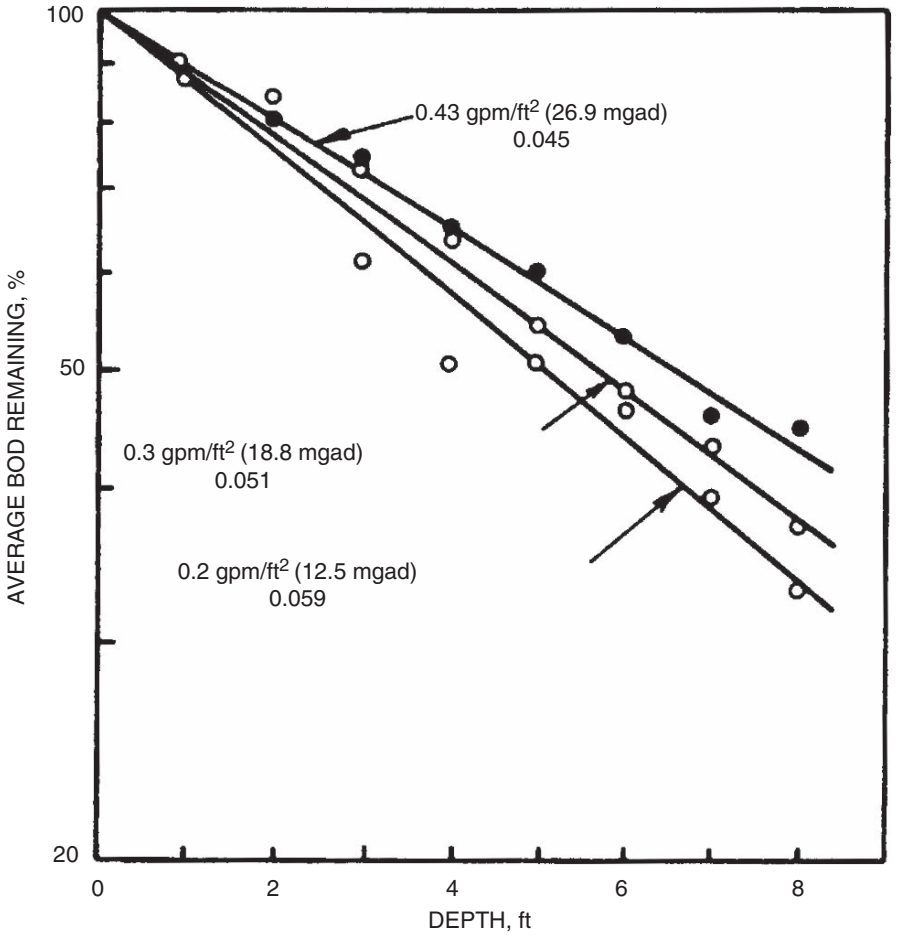


Fig. 9.11. Relation between filter depth and percent BOD Remaining at various hydraulic loads (*Source*: Water and Waste Engineering; Copyright and Published By Reuben H. Donnelley Corp.).

of slope in Figure 9.11 is rewritten as

$$\ln(-\text{slope}) = \ln(K) - n\ln(q) \tag{47}$$

Accordingly the slopes of the BOD_5 remaining versus depth curves (Figure 9.11) can be plotted against their respective hydraulic rates on a logarithmic graphical sheet and the constant n is determined to be 0.39 shown in Figure 9.12.

Finally Equation (46) is reconsidered and the constant K can be determined as illustrated in Figure 12, by plotting $Q^{-n}D^{m}$ vs $100L_e/L_o$ on semilogarithmic sheet. The K value at the wastewater temperature, 14°C, has been determined to be 0.375. The K rate at 20°C is then calculated by Equation (25) assuming $\theta = 1.035$ (117):

$$K \text{ at } 20°\text{C} = 0.375 \times 1.035^{(14-20)} = 0.46$$

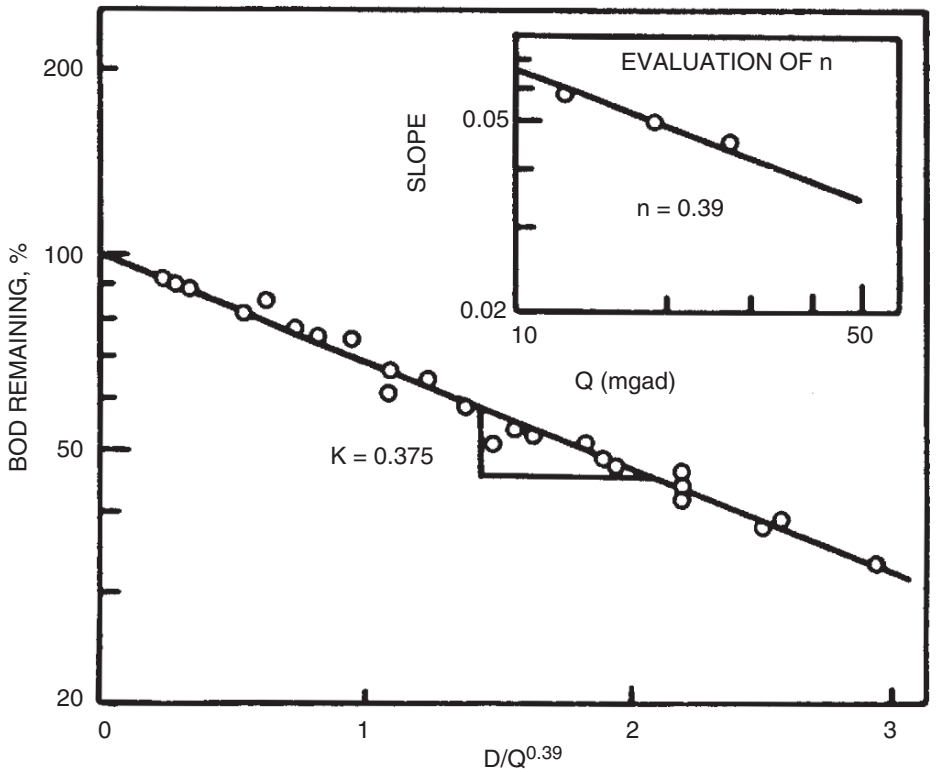


Fig. 9.12. Diagrams for the determination of constants n and K in Eckenfelder's model (*Source*: *Water and Waste Engineering,* Copyright and Published by Redben H. Donnelicy Corp.).

Substituting values of $m = 1$, $K = 0.46$ and $n = 0.37$ in Equations (30) and (45), the BOD removal relationships for the specific filter medium tested, the specific wastewater treated, and under the described operational conditions are:

$$L_e/L_o = \exp(-0.46\ D/q^{0.39}) \tag{48}$$

for a trickling filter system without recirculation, and

$$\frac{L_e}{L_o} = \frac{\exp(-0.46D/q^{0.39})}{(1+R) - R[\exp(-0.46D/q^{0.39})]} \tag{49}$$

for a trickling filter system with recirculation.

11.7. Example 7

Describe the mathematical and experimental approaches for confirmation or determination of the constants in the general design model, Equation (41).

Solution

Equation (41) can be rewritten as

$$L_e/L_o = \exp[-K'(Dq^{-1})^{0.5}] \tag{50}$$

or,

$$L_e/L_o = \exp[-K'(Dq^{-n'})^{m'}] \tag{50}$$

Where K', n', and m' are constants, and L_e, L_o, D, and q have been defined earlier, It can be seen that the mathematical and experimental approaches used in Example 6 can also be applied to the general design model for determining the constants.

11.8. Example 8

A trickling filter plant, shown in Figure 9.13, consists of a primary sedimentation basin, a circular trickling filter 80 ft in diameter with an 8 ft depth of 3/4-in Raschig rings as the media, and a secondary sedimentation basin. The following are some given operational conditions and design criteria:

(a) Characteristics of 3/4-in. Rasching rings:
  Surface area = 75.8 ft^2/ft^3
  Reaction-rate constant, $K = 0.55$ at 20°C
  Packing constant, $n = 0.7$
  Another packing constant, $m = 1.0$
(b) Raw-wastewater flow = 0.5 MGD
(c) Wastewater temperature = 25°C (both influent and effluent)
(d) BOD_5 concentration of raw wastewater = 210 mg/L
(e) Indirect recirculation to the wet well = 0.3 MGD
(f) Direct recirculation around the trickling filter = 0.4 MGD
(g) Overflow rate of the primary sedimentation basin = 500 gpd/ft^2

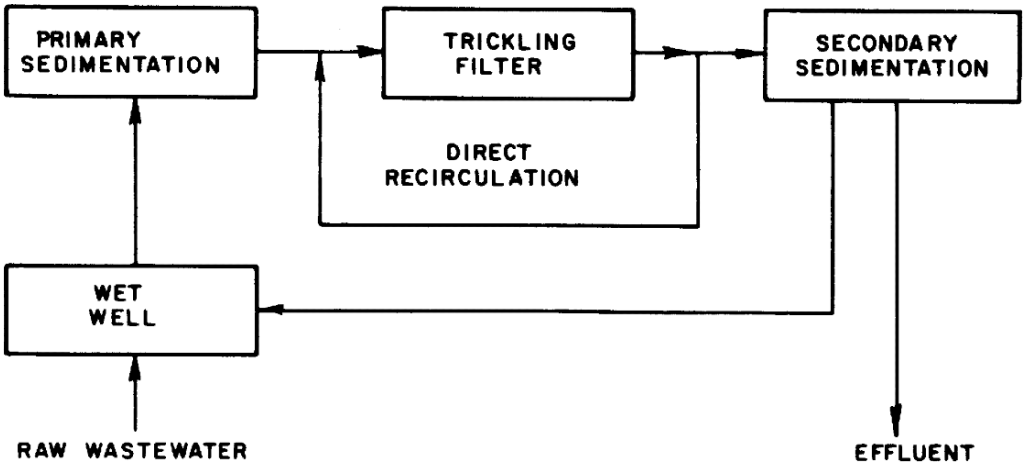


Fig. 9.13. Single-stage trickling filter with direct and indirect recirculation.

(h) Temperature coefficient, $\theta = 1.035$
(i) Use Eckenfelder's Model.

Determine the following design and operational parameters:

(a) BOD_5 removal efficiency of primary clarification
(b) BOD_5 removal efficiency of the trickling filter
(c) Effluent BOD_5 concentration, mg/L

Solution

(a) Percent BOD_5 removal of primary clarification

$$= 36.5\% \text{ based on an overflow rate of } 500\,\text{gpd/ft}^2$$

(b) Area

$$A = (80/2)^2(3.14) = 5{,}024\,\text{ft}^2 = 0.115\,\text{ac.}$$

$$\text{Wastewater flow, } Q = 0.5\,\text{MGD}$$

Trickling filter's hydraulic loading excluding recirculation

$$q = Q/A = 0.5/0.115 = 4.35\,\text{mgad}$$

Direct recirculation flow

$$Q_{\text{rd}} = 0.4\,\text{MGD}$$

Indirect recirculation flow

$$Q_{\text{ri}} = 0.3\,\text{MGD}$$

Recirculation ratio

$$R = (Q_{\text{rd}} + Q_{\text{ri}})/Q$$
$$= (0.4 + 0.3)/0.5 = 1.4$$

$n = 0.7$ which will not be affected by temperature
K at 20°C = 0.55

$$K \text{ at } 25^\circ\text{C} = (K \text{ at } 20^\circ\text{C})\theta^{(25-20)} \tag{25}$$
$$= 0.55 \times 1.035^5$$
$$= 0.653$$

$$L_e/L_o = \frac{\exp(-KD^m/q^n)}{(1+R) - R[\exp(-KD^m q^n)]} \tag{45}$$
$$= \frac{\exp(-0.653 \times 8/4.35^{0.7})}{(1+1.4) - 1.4[\exp(-0.653 \times 8/4.35^{0.7})]}$$
$$= 0.071$$

Percent BOD_5 removal of trickling filter

$$= 1 - 0.071 = 93\%$$

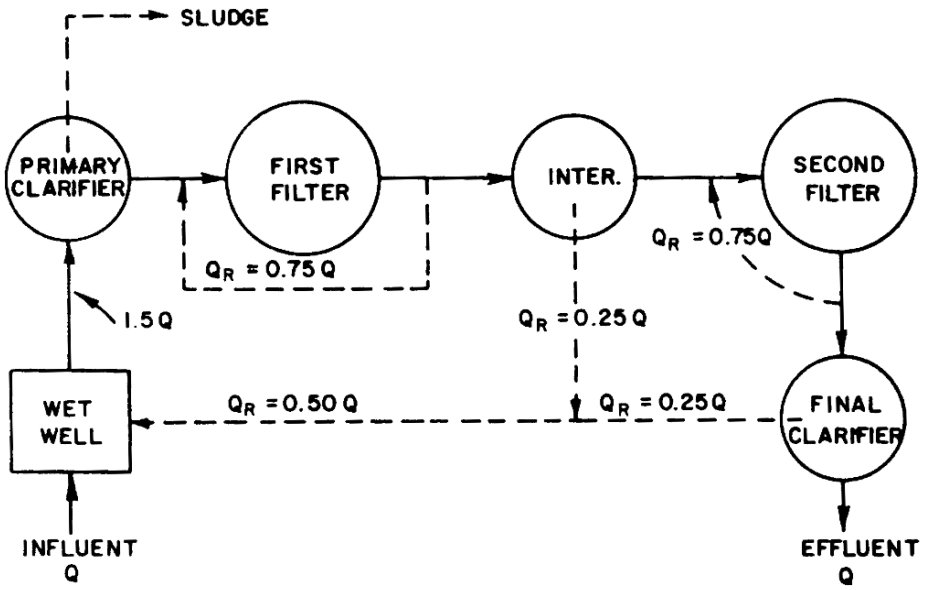


Fig. 9.14. Flow scheme of a two-stage trickling-filler plant (*Source*: Harper & Row Publishers, NY).

(c) Effluent BOD_5

$$BOD_{effluent} = (\text{Influent } BOD_5)(1 - \%\text{removal of primary clarifier})(L_e/L_o)$$
$$= (210\,\text{mg/L})(1 - 0.365)(0.071) = 9.47\,\text{mg/L}$$

11.9. Example 9

A new two-stage rock media trickling filter plant shown in Figure 9.14 is to be designed using the National Research Council formula. The design criteria and other given data are presented below:

(a) Wastewater flow, $Q = 2.4$ MGD
(b) BOD_5 concentration of raw wastewater (L_r) = 400 mg/L
(c) Design criteria of primary clarifier:
   1. Overflow rate equals to 500 gpd/ft^2 based on Q, or equals to 750 gpd/ft^2 based on ($Q + Q_r$)
   2. Depth = 7 ft minimum
   3. BOD_5 removal efficiency of primary clarifier = 25%
   4. Number of clarifiers = 2
   5. Shape = circular
(d) Design criteria of trickling filters:
   1. BOD_5 loading = 50 lb/1000 ft^3/d
   2. Hydraulic loading = 10–30 mgad
   3. Depth = at least 6 ft
   4. Shape = circular
   5. Number of filters = 2
   6. Volume of filter media = total volume divided equally between the first-stage and second-stage filters

(e) Design criteria of intermediate clarifiers
   1. Overflow rate = 1000 gpd/ft^2
   2. Depth = 7 ft minimum
   3. Number of clarifiers = 2
   4. Shape = circular

(f) Design criteria of final clarifiers
   1. Overflow rate = 800 gpd/ft^2
   2. Depth = 7 ft minimum
   3. Number of clarifiers = at least 2
   4. Shape = circular

Determine the following design and operational parameters:

(a) Diameter and depth of primary clarifiers
(b) Diameter, depth, and hydraulic loading of first-stage trickling filters
(c) Diameter and depth of intermediate clarifiers
(d) Diameter and depth of second-stage trickling filters
(e) Diameter and depth of final clarifiers
(f) BOD_5 loadings to the first-stage and second-stage filters
(g) BOD_5 concentration in the plant effluent

Solution

(a) $Q = 2.4$ MGD
Area of each primary clarifier based on Q

$$A = [(2.4 \times 10^6)/2]\ \text{gpd}/(500\ \text{gpd/ft}^2)$$
$$= 2400\ \text{ft}^2$$

Area of each primary clarifier based on $(Q + Q_r)$

$$A = [(1.5 \times 2.4 \times 10^6)/2]\text{gpd}/(750\ \text{gpd/ft}^2)$$
$$= 2400\ \text{ft}^2$$

Provide a side-wall depth of 8 ft plus freeboard
Diameter of each primary clarifier

$$d = (2400\ \text{ft}^2 \times 4/3.14)^{0.5}$$
$$= 55.28\ \text{ft, use } 55.5\ \text{ft}$$

(b) Volume of each trickling filter

$$= (400)(1 - 0.35)(8.34)(2.4/4)(50/1000)$$
$$= 26{,}021\ \text{ft}^3 \text{ (two first-stage and two second-stage filters)}$$

Depth = 6 ft (trial value)
Area of each trickling filter

$$A = (26{,}021/6)$$
$$= 4336.8\ \text{ft}^2$$
$$= 0.1\ \text{ac.}$$

Hydraulic loading to each filter

$$= (1.5 + 0.75)(2.4/2)/0.1$$
$$= 27.0\,\text{mgad}$$

Diameter

$$d = (4{,}336.8 \times 4/3.14)^{0.5}$$
$$= 74.3\,\text{ft, use } 74.5\,\text{ft}$$

(c) Area of each intermediate clarifier

$$A = (1.5 - 0.25)[(2.4 \times 10^6)/2]\text{gpd}(1000\,\text{gpd/ft}^2)$$
$$= 1500\,\text{ft}^2$$

Use side wall depth of 7 ft
Diameter of each intermediate clarifier

$$d = (1500\,\text{ft}^2 \times 4/3.14)^{0.5}$$
$$= 43.7\,\text{ft(use } 44\,\text{ft)}$$

(d) Diameter and depth of second-stage trickling filters are identical to that of first-stage trickling filters.

(e) Area of each final clarifier

$$A = (1.25 - 0.25)[(2.4 \times 10^6)/2]\text{gpd}/(800\,\text{gpd/ft}^2)$$
$$= 1500\,\text{ft}^2$$

Use side wall depth of 7 ft
Diameter = 43.7 ft, use 44 ft

(f) BOD_5 removal efficiency of primary clarifiers = 35%
BOD_5 loading of first-stage filters

$$(W/V) = \frac{(1 - 0.35)(400)(2.4/2)(8.34)}{26.021 \times 1{,}000}$$
$$= 100\,\text{lb/d}/1000\,\text{ft}^3$$
$$= 4{,}356\,\text{lb/d/ac.-ft.}$$

For first-stage trickling filters

$$W = (1 - 0.35)(400)(2.4/2)(8.34) = 2{,}602.1\,\text{lb/d}$$
$$R = Q_r/q = (0.5Q + 0.75Q)/Q = 1.25$$
$$F = (1 + R)/(1 + 0.1R)^2$$
$$= (1 + 1.25)/(1 + 0.1 \times 1.25)^2 = 1.78$$
$$V = \text{Area} \times \text{depth} = 0.1\,\text{ac.} \times 6\,\text{ft} = 0.6\,\text{ac.-ft}$$

$$E_1 = 100[1 + K_{\text{nrc}}(W/VF)^{0.5}]^{-1} \tag{12}$$
$$E_1 = 100[1 + 0.0085(2602.1/0.6 \times 1.78)^{0.5}]^{-1}$$
$$E_1 = 100[1 + 0.0085(4336.8/1.78)^{0.5}]^{-1} = 70(\%)$$

For second-stage trickling filters:

$$W' = (1 - E_1)(W) = (1 - 0.7)(2602.1) = 780.63$$
$$R = Q_r/Q = (0.25Q + 0.75Q)/Q = 1$$
$$F = (1 + 1)/(1 + 0.1 \times 1)^2 = 1.65$$
$$V = 0.6\,\text{ac.-ft}$$
$$E_2 = 100[1 + 0.0085(W'/VF)^{0.5}/(1 - E_1)]^{-1} \tag{16}$$
$$E_2 = 100[1 + 0.0085(780.63/0.6 \times 1.65)^{0.5}/(1 - 0.7)]^{-1} = 56(\%)$$

(g) Plant efficiency E

$$E = 100 - 100[(1 - 0.35)(1 - 0.70)(1 - 0.56)] = 92\%$$

Estimated effluent BOD_5 concentration

$$= (400\,\text{mg/L})(1 - 0.92)$$
$$= 32\,\text{mg/L}$$

11.10. Example 10

Figure 9.15 shows the optimized design curves for the Eckenfelder model for an influent of 1 MGD (4,000 m^3/d) at 18°C and an application rate of 30 mgad (29.65 m^3/m^2/d). The removal efficiency is plotted on the abscissa, and the variables, filter depth, recirculation, and radius, which determine the trickling filter area, are plotted on the ordinate beginning with the depth that contributed least to the cost of the filter (37). Only one curve is developed because in this model, for any given BOD_5 removal efficiency, the filter design is related only to filter depth and recirculation. Based on the concept that BOD_5 removal is related only to wastewater flow and the loading has no effect on BOD_5 removal. The BOD_5 removal efficiency increases with depth up to the maximum desirable depth, then recirculation to a ratio of four will further increase removal efficiency after which the filter area and volume must be increased, thus decreasing wastewater flow.

Figure 9.16 shows the maximum recirculation ratios for various plant influent rates for varying filter radii. This figure can be used in conjunction with any trickling filter design model.

Design a trickling filter plant graphically using the optimized Eckenfelder model (Figure 9.15) and considering the following environmental conditions:

(a) Wastewater influent flow = 10 MGD
(b) BOD_5 concentration to the filters = 150 mg/L
(c) Wastewater temperature = 18°C
(d) Desired effluent BOD_5 = 25 mg/L
(e) Desired filter depth = shallow, less than 10 ft.

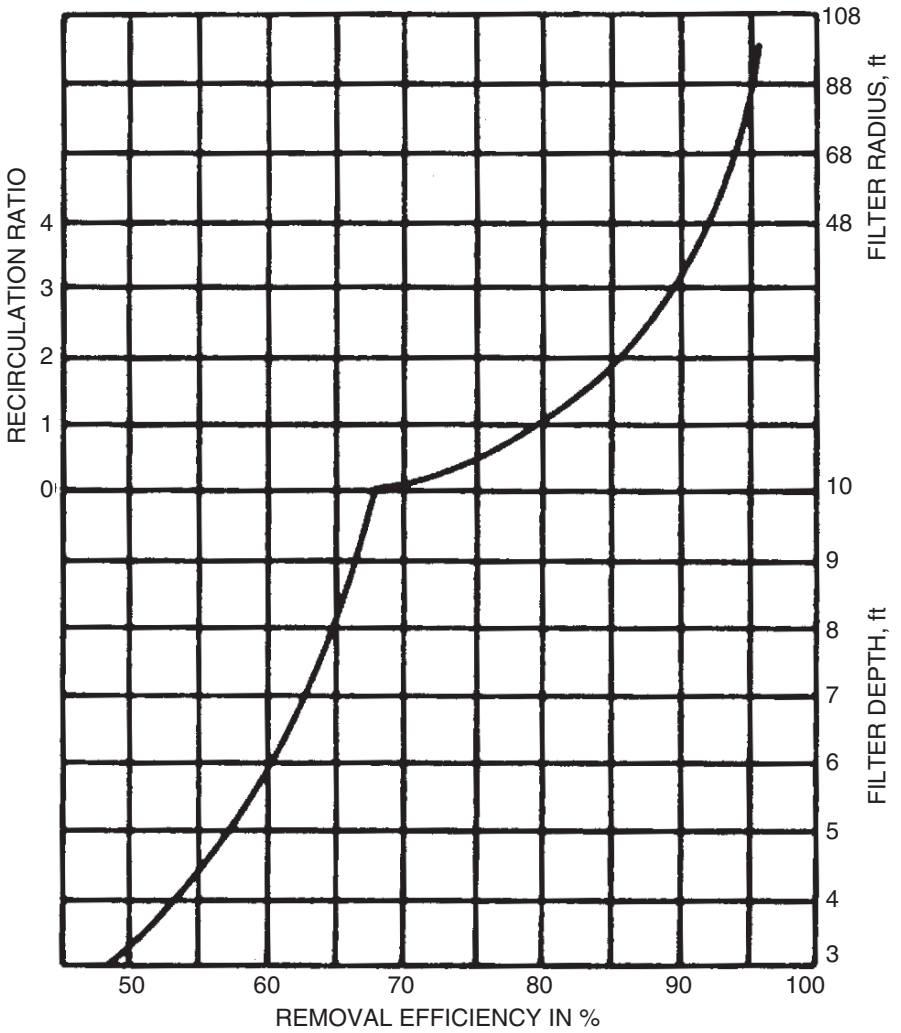


Fig. 9.15. Trickling filter design characteristics at 30 mgad and 18°C maximum depth 10 ft; plant influent 1 MGD Eckenfelder model (*Source*: ASCE).

Solution

BOD_5 removal efficiency $= (150 - 25)/150 = 0.833$

Using Figure 9.15, the curve for BOD_5 removal efficiency shows that a filter depth of 10 ft and a recirculation rate of 1.25 is required for 83.3% BOD_5 removal by the filter.

Because the recirculation rate is less than 4.0, dividing the total flow, $10(1 + 1.25)$ MGD, by 30 mgad to determine the filter area, or by finding on the 10-MGD influent curve in Figure 9.16 the point where the recirculation rate is 1.25, a radius of 104 ft is required

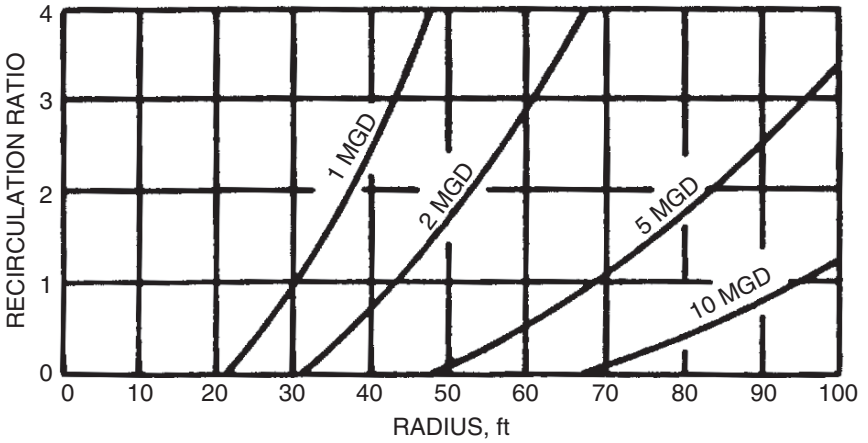


Fig. 9.16. Maximum recirculation ratios for various plant influent rates for varying filter radii (*Source*: ASCE).

Because the radius of one filter is over 100 ft and it is desirable for flexibility in plant operation to use at least two filters, each filter treating 5 MGD with a recirculation ratio of 1.25, and a depth of 10 ft, must have a radius of 72 ft. The hydraulic loading is 30 mgad.

11.11. Example 11

Figure 9.17 present optimized curves for the Galler-Gotaas model (37) for effluent BOD_5 concentrations related to influent BOD_5 concentrations of 100 to 400 mg/L, where the influent BOD_5 is the abscissa and the depth, recirculation, and radius are plotted on the ordinate.

Design a trickling filter plant graphically using the optimized curves of the Galler-Gotaas model and considering the same environmental conditions for Example 10.

Solution

Using Figure 9.17, find the point on the 150 mg/L influent loading curve where the effluent BOD_5 is 25 mg/L. At this point the depth is 10 ft and the recirculation ratio is 0.7.

By dividing the total flow, $10(1 + 0.7)$ MGD, by 30 mgad to determine the filter area, or by finding the 10 MGD influent curve and the point where the recirculation ratio is 0.7 in Figure 9.16, a radius of 88 ft for one filter, or two filters each having an influent rate of 5 MGD and a radius of 63 ft.

Thus, the design has two trickling filters having a depth of 10 ft, a radius of 63 ft. a recirculation ratio of 0.7 and a hydraulic load of 30 mgad.

11.12. Example 12

Figure 9.18 presents the Galler-Gotaas model for the filter depth range 10 ft to a maximum of 20 ft. A family of influent BOD_5 curves indicates that efficiency is a function of the organic loading raised to a power.

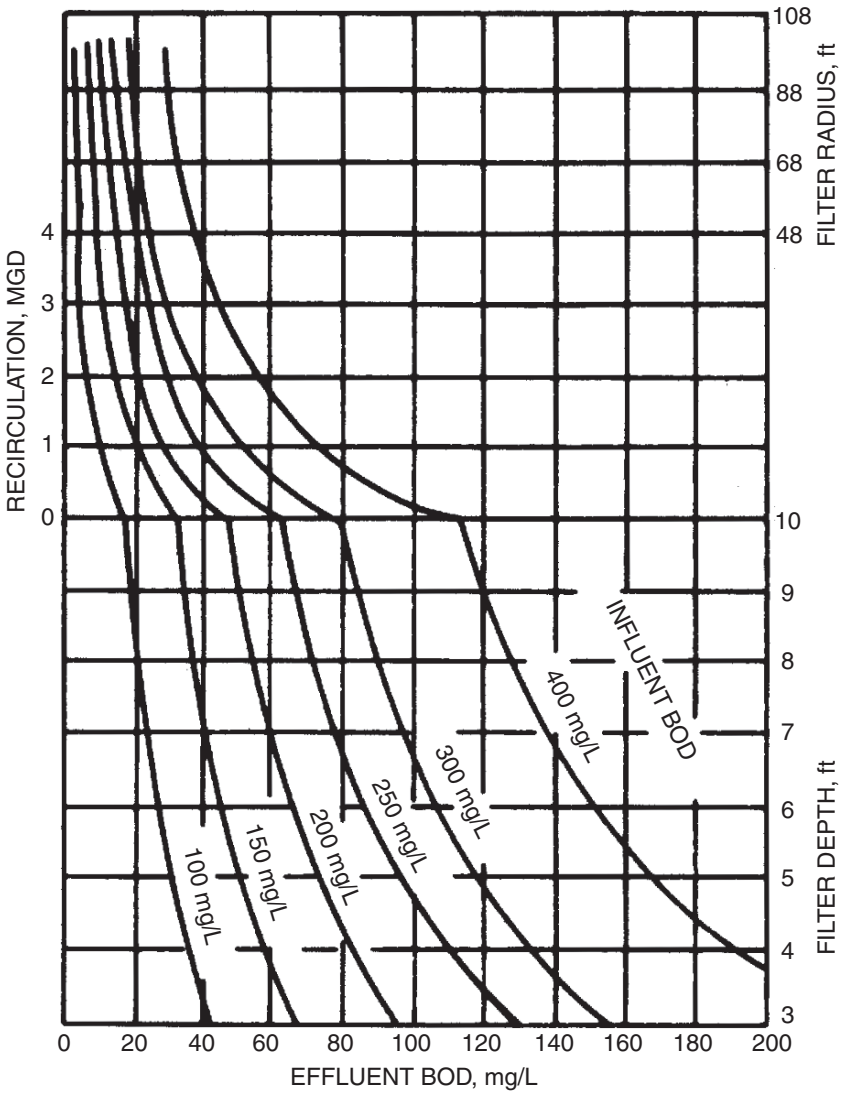


Fig. 9.17. Trickling filter design characteristic at 30 mgad and 18°C maximum depth, 10 ft; plant influent, 1 MGD Galler-Gotaas model (*Source*: ASCE).

Discuss the applicability of the figure and the advantages and disadvantages of deep depth trickling filters.

Solution

Ten feet has often been considered to be a maximum allowable depth for conventional trickling filters, assuming adequate oxygen supply. When the depth constraint is increased as shown in Figure 9.18, the most economical filter depth reaches the maximum depth

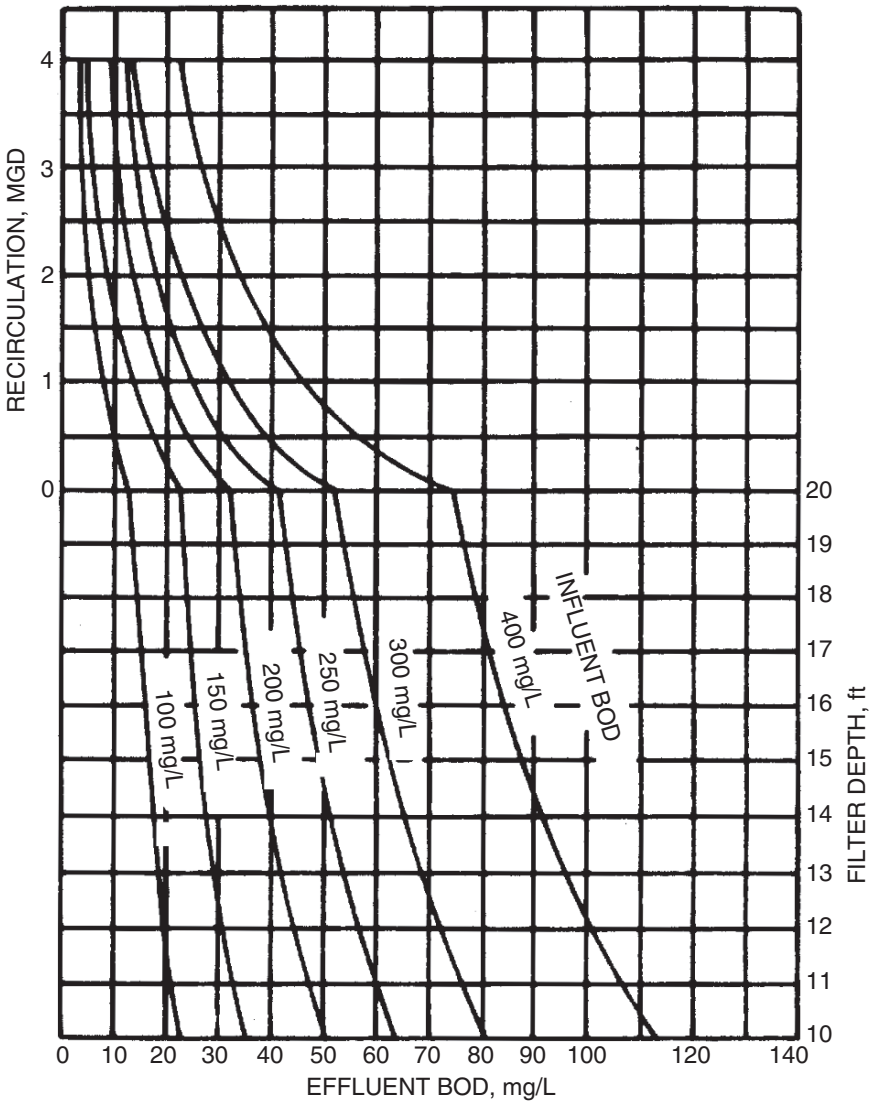


Fig. 9.18. Trickling filter design characteristic at 30 mgad and 18°C maximum depth, 20 ft; plant influent, 1 MGD Galler-Gotaas model (*Source*: ASCE).

when the hydraulic gradient of the treatment plant is such that pumping the influent to a deep filter is not required, and sufficient ventilation for the necessary oxygen supply is available.

If it is necessary to cover the filters and provide forced ventilation with odor and bacterial control of the existing air, deeper filters are more economical. If pumping is required, and covering the filter is unnecessary, the maximum depth for which an adequate oxygen supply can be provided is the most economical.

Analysis of the use of forced air ventilation with blowers at a rate of 2 ft^3/gal of influent wastewater showed (37) that the filter variables follow the curves plotted in Figure 9.18. If the recirculation required for a shallow filter is greater than 0.5 times the influent rate, a deeper filter with air blowers is more economical than the shallow filter.

11.13. Example 13

Figure 9.19 presents optimized curves for the National Research Council (NRC) model, for an influent of 1 MGD (4000 m^3/d) at 18°C, an application rate of 30 mgad (29.65 m^3/m^2/d) and a maximum depth of 10 ft.

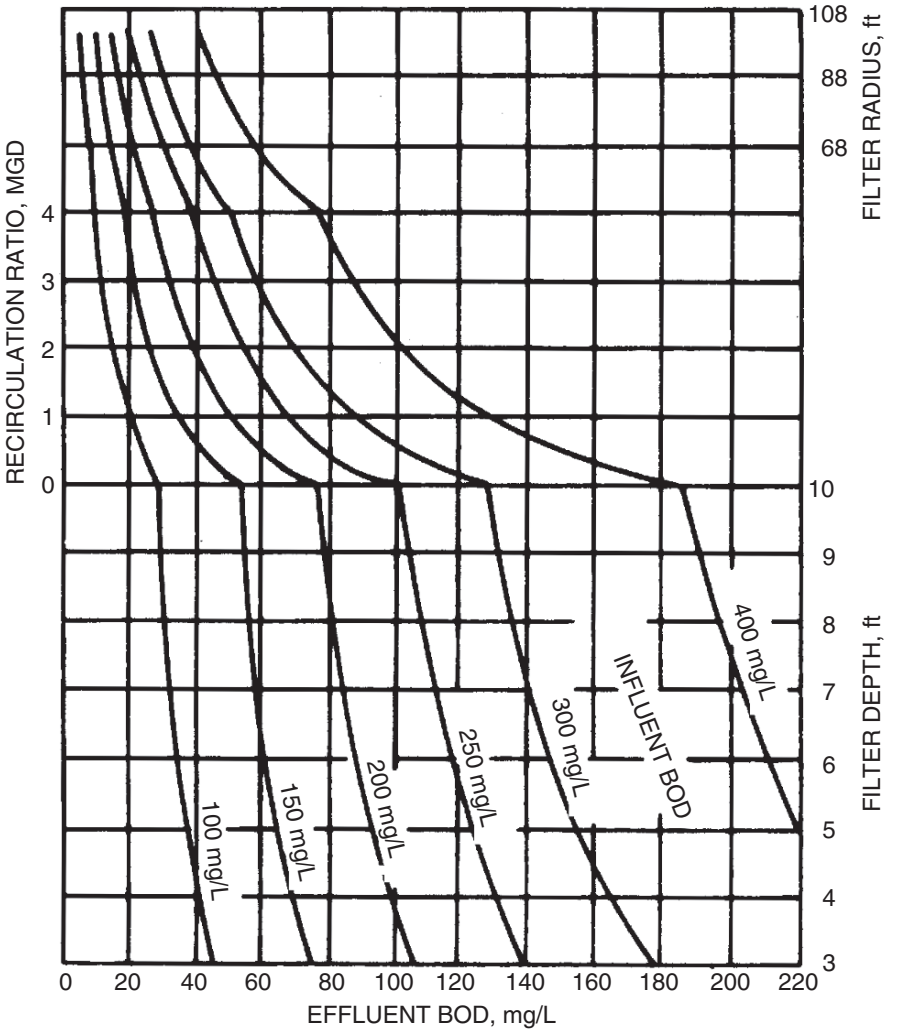


Fig. 9.19. Trickling filter design characteristic at 30 mgad and 18°C: maximum depth, 10 ft; plant influent, 1 MGD, NRC model.

Design a trickling filter plant graphically using the optimized design curves of the NRC model, and considering the same environmental conditions for Example 10.

Solution

The BOD_5 removal increases as the filter depth is increased. In Figure 9.19 the 150 mg/L influent BOD_5 concentration curves reaches an effluent BOD_5 of 25 mg/L when the depth is 10 ft.

Once the maximum depth constraint is reached, the depth of the filter remains constant at maximum and recirculation is now necessary to improve BOD_5 removal and increases to a maximum of four volumes per volume of influent wastewater. From Figure 9.19, the required recirculation ratio is 2.1 for an effluent BOD_5 concentration of 25 mg/L.

The recirculation ratio then remains constant at the determined value (note: less than 4.0) until the maximum radius is reached, thus reducing the hydraulic flow rate.

In Figure 9.16, it is seen that the filter radius is over 100 ft when the recirculation ratio is 2.1, thus, two trickling filters are required. Using two filters, each treating 5 MGD as shown the radius of each filter is 82 ft.

Therefore, the design using the NRC model has two filters having a depth of 10 ft, a radius of 82 ft. and a recirculation ratio of 2.1, with a hydraulic loading of 30 mgad.

11.14. Example 14

The Upper Mississippi River-Great Lakes Board (UMRGLB) Model (33) is graphically shown in Figure 9.20. Only one curve is developed because the effluent BOD_5 is a function of the influent BOD_5 raised to unit power. The trickling filter depth can be at the minimum allowed because there is no interaction between BOD_5 and filter depth in the UMRGLB model. The volume of the filter is established by the maximum loading criteria of 110 lb $BOD_5/1000\,ft^3/d$(1.77 $kg/m^3/d$) and the area by the hydraulic loading of 30 mgad (29.65 $m^3/m^2/d$) for a filter depth of less than 7 ft and greater than 5 ft. A constraint of a maximum recirculation rate of four should be placed on this model. The hydraulic loading should be between 10 and 30 mgad. The filter influent BOD_5 concentration should not be greater than three times that of the effluent.

Design a trickling filter plant graphically using the optimized design curves of the UMRGLB model, and considering the same environmental conditions for Example 10.

Solution

BOD_5 removal efficiency $= (150 - 25)/150 = 0.833$

Recirculation ratio $= 1.5$ based on the BOD_5 removal efficiency in Figure 9.20.

Analysis shows that the model requirements of a maximum hydraulic loading of 30 mgad and minimum depth of 5 ft determine the size of filter. For a hydraulic loading of 30 mgad, and a recirculation ratio of 1.5, the filter area (A) required for 10 MGD is:

$$A = 10(1 + 1.5)\,\text{MGD}/30\,\text{mgad}$$

$$= 0.833\,\text{ac.} = 364{,}000\,\text{ft}^2$$

Considering two filters, the radius of each will be 76 ft.

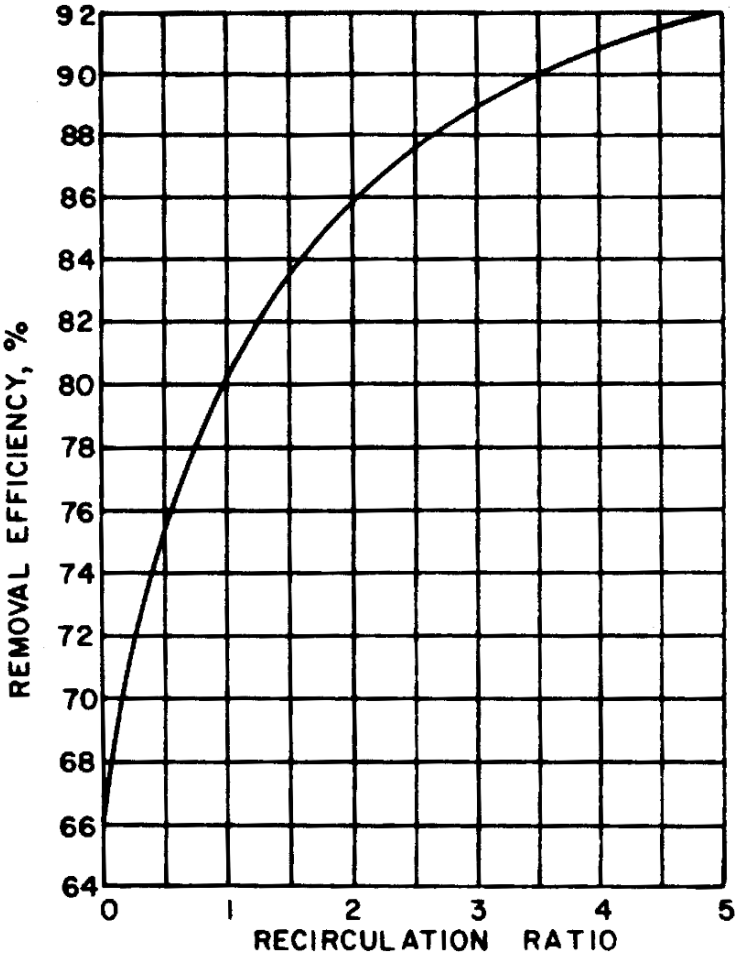


Fig. 9.20. Trickling filter design characteristic at 30 mgad and 18°C maximum depth, 10 ft; plant influent, 1 MGD Upper Mississippi River-Great Lakes model.

Because the area and radius are determined by the recirculation rate and 30 mgad hydraulic loading required, the minimum allowable filter depth of 5 ft to minimize the filter volume can be used. Now the BOD_5 loading should be checked. The BOD_5 loading is:

$$W = (10)(8.34)(150) = 12{,}500\,\text{lb/d}$$

or

$$W/V = 12{,}500(1{,}000)/(2 \times 3.14 \times 76^2 \times 5) = 76\,\text{lb/d}/1000\,\text{ft}^3$$

which is less than $110\,\text{lb/d}/1000\,\text{ft}^3$.

Therefore, the design using the UMRGLB model requires two filters each having a depth of 5 ft, a radius of 76 ft, and a recirculation ratio of 1.5 with a hydraulic loading of 30 mgad.

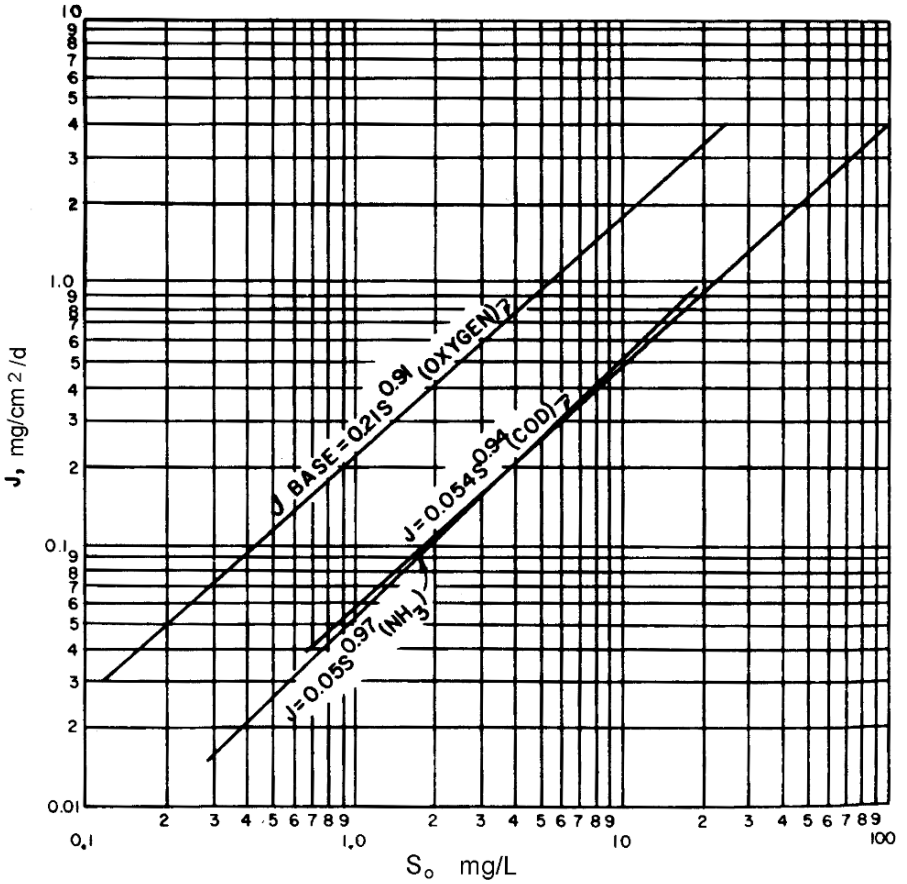


Fig. 9.21. Biofilm model base computations (*Source*: US EPA).

11.15. Example 15

Figures 9.21 and 9.22 have been developed from the Biofilm model and illustrate that over a wide range of substrate concentrations, especially that of interest in municipal wastewater treatment. Equation (34) is a suitable approximation of the biofilm model over a limited substrate range (1 to 100 mg/L) as indicated in Figure 9.22 for the temperature range of typical wastewater. The coefficient p represents the slope of the curves and is 0.94 for the curves given. The coefficient k_T represents the flux per unit area when S_o is 1.0 mg/L. Assume the wastewater to be treated can be characterized by the empirical formula of $C_{10}H_{19}O_3N$ and has a BOD_L concentration of 20 mg/L in the surrounding biofilm.

Determine:

(a) The flux of substrate into the biofilm
(b) The flux of oxygen required for this biochemical oxidation; and
(c) The required oxygen concentration.

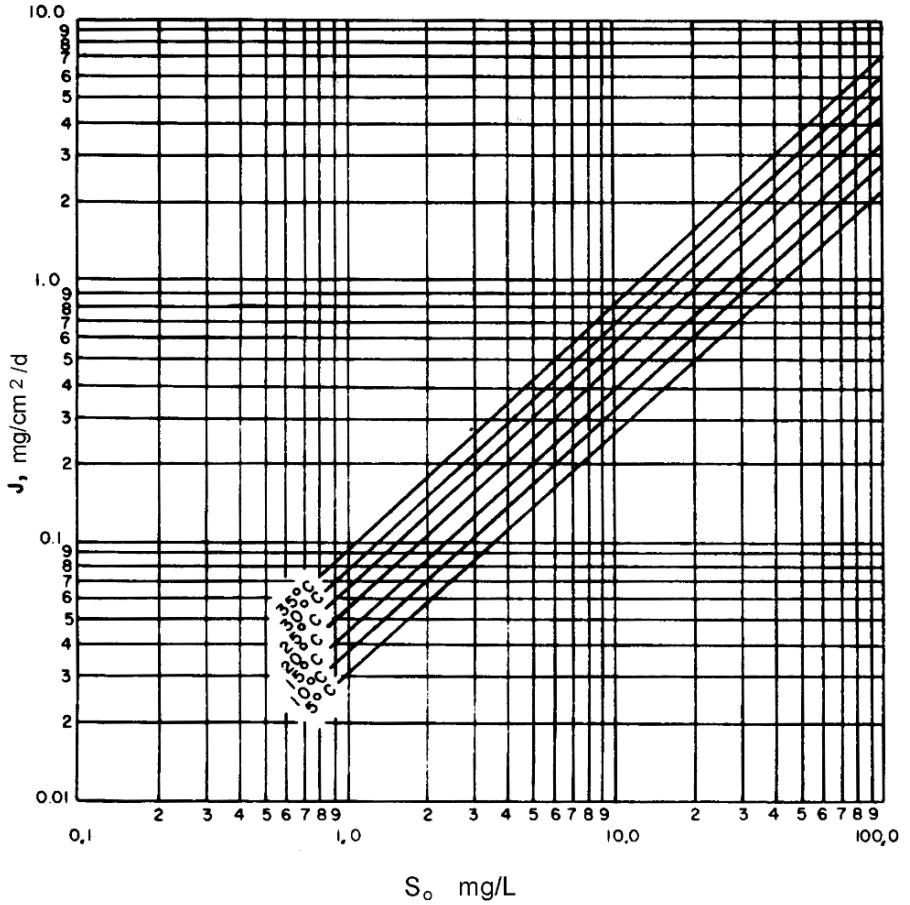


Fig. 9.22. Predicted effects of temperature variations on J vs S (*Source*: US EPA).

Solution

(a) Two moles of $C_{10}H_{19}O_3N$ would have a theoretical oxygen demand or ultimate BOD_L of 800 g [according to Equations (1) to (3)] requiring 10 mol or 320 g of oxygen for oxidation and synthesis into bacterial cells. Thus 0.4 mg of oxygen would be needed to diffuse into the biofilm of each mg of BOD_L that diffuses in and is consumed by the bacteria. Figure 9.21 indicates the flux of substrate into the biofilm (J) would be 0.9 mg/cm^2/d if organics (S_o) were substrate limiting at 20 mg/L.

(b) The flux of oxygen required for this to be the case would be:

$$\text{Flux} = (0.9\,\text{mg BOD}_L/\text{cm}^2/\text{d})(0.4\,\text{mg O}_2/\text{mg BOD}_L)$$
$$= 0.36\,\text{mg O}_2/\text{cm}^2/\text{d}$$

(c) Figure 9.21 indicates that the required oxygen concentration (S_0) would be 1.8 mg/L when the flux of oxygen (J) equals to 0.36 mg O_2/cm^2/d. Dissolved oxygen would limit the rate of the reaction if the minimum oxygen concentration were lower than this value. Figure 9.22 is used only for temperature adjustments.

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NOMENCLATURE

a = media specific surface area per unit volume, L^2/L^3
A = surface area of the trickling filter, L^2
A_c = biofilm area, L^2
C = a constant
d = diameter of the trickling filter, L
D = depth of the trickling filter, L
D_c = diffusion coefficient within the biofilm, L^2/T
D_w = diffusion coefficient of the chemical species through water, L^2/T
E = fractional efficiency of BOD removal
E_n = energy consumption, kwh/yr
E_T = efficiency at temperature, T°C
E_1 = percent efficiency of BOD removal of single-stage trickling filter, calculated by the NRC model, %
E_2 = percent efficiency of BOD removal of second-stage trickling filter, calculated by the NRC model, %
E_{20} = efficiency at 20°C
F = recirculation factor
F_5 = sludge production factor, M solids/ M BOD_5
H = discharge head, L
J = flux of substrate into the biofilm, $M/L^2/T$
J_o = surface flux of the chemical species, M/T
K = maximum use rate of the rate-limiting substrate, M/T/M
k_T = reaction rate at temperature T
k' = a constant
k'' = a constant
K = a constant
K_{ace} = coefficient used in the model developed by the US Army Corps of Engineers
K_e = rate of BOD removal, base e
K_{nrc} = coefficient used in the National Research Council Model for trickling filter design = 0.0085
K_p = performance measurement parameter
K_{10} = rate of BOD removal, base 10
L = applied removable BOD, M/L^3
L_a = applied BOD after dilution by recirculation, M/L^3
L_D = BOD remaining in the effluent at depth 0. M/L^3
L_e = BOD of effluent, M/L^3
L_o = BOD of influent, M/L^3

m = a constant
n = a constant
p = a coefficient
P_s = sludge produced, M/T
q = hydraulic loading, $L^3/L^2/T$
q_w = wetting rate (surface application rate), $L^3/T/L^2$
Q = influent wastewater flow rate, L^3/T
Q_{ave} = average daily flow, L^3/T
Q_{pd} = peak daily flow, L^3/T
Q_{pm} = peak monthly flow, L^3/T
Q_r = recirculation flow rate, L^3/T
R = recirculation ratio
S = substrate concentration, M/L^3
S_c = substrate concentration within the biofilm cellular matrix, M/L^3
S_e = effluent substrate concentration, M/L^3
S_i = Interior substrate concentration, M/L^3
S_m = substrate mass, M
S_0 = bulk liquid substrate concentration, M/L^3
S_s = substrate concentration at the biofilm surface, M/L^3
t = time
T = temperature, °F or °C
V = volume of attached growth media, L^3
W = BOD loading to a single-stage filter, M/T
W' = BOD loading to a second-stage filter, M/T
X_c = bacterial concentration within the biofilm, M/L^3
X_v = volatile biological solids concentration, M/L^3
Y = depth of a stagnant liquid layer outside the slime-liquid interface, L
Y_c = depth of a biofilm within the cellular matrix, L
Y_e = effective depth of biofilm at which $S_c = S_i$, L
Z = direction z
θ = temperature coefficient

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10
Rotating Biological Contactors

Lawrence K. Wang, Zucheng Wu, and Nazih K. Shammas

CONTENTS

Abstract Rotating biological contactor (RBC) is an attached-growth biological process, which consists of a series of rotating plastic media all coated with a layer of biofilm. The biofilm or slime on the media aerobically react with substances in a waste stream for bio-oxidation and nitrification, or anaerobically react with the substances for denitrification. This chapter discusses the theory, performance, design procedures, process control, applications, limitations, environmental impact, and design considerations of RBC process.

Key Words Rotating biological contactor • RBC • attached-growth • biological system • bio-oxidation • nitrification • denitrification • environmental impact • design examples.

1. INTRODUCTION

Rotating biological contactors (RBCs) were originally developed in Europe and recently accepted by America and Asia. The process system, as shown in Figure 10.1 is primarily a fixed-film biological reactor consisting of a synthetic medium mounted on a horizontal

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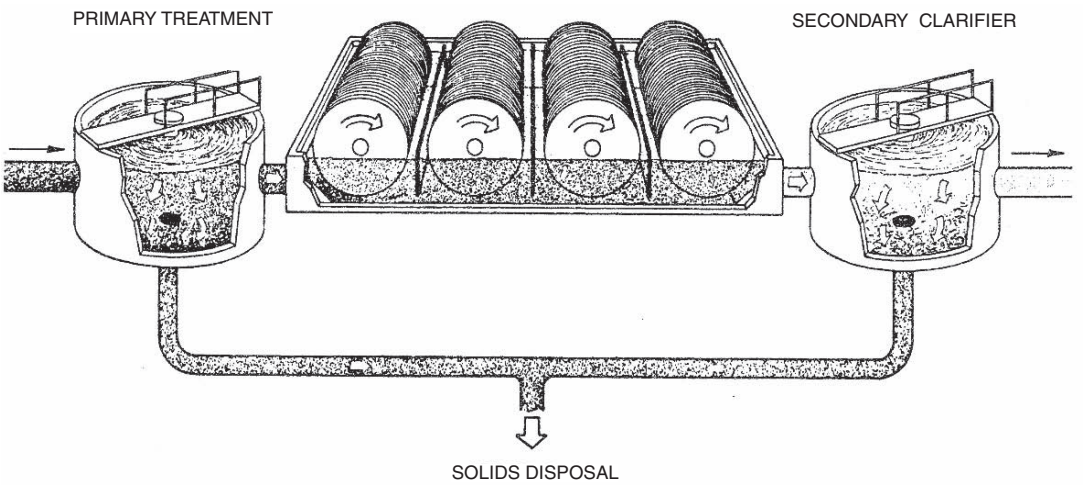

Fig. 10.1. Typical RBC process (*Source*: US EPA).

shaft and placed in a contour-bottomed tank. The general concept of rotating biological contactors is to let wastewater flow through the tank, and to rotate the medium in the wastewater to be treated, alternatively exposing the medium (and the attached biological growth) to air and the wastewater. The slowly rotated media are about 40% immersed in the wastewater for aerobic removal of organic waste by the biological film developing on the media. The lattice-structured medium, and to a lesser extent the disc structure, is fragile and should be protected from direct exposure to wind, sun, and weather fluctuation. Therefore, the media are usually enclosed in a superstructure or individual shaft covers.

Media rotation can be provided by either mechanical drives or air-motivated rotation. Rotation not only results in exposure of the film to the atmosphere as a means of aeration, but also provides rotational shear forces for stripping off the excess biomass on the medium. The stripped biological solids are maintained in suspension by the mechanical mixing action of the rotation medium, or by supplemental diffused air, depending on the driving force of rotation. The air-driven system, in rotating the media by diffused air generated near the tank bottom, alleviates the development of undesirable anaerobic Conditions, and also reduces the oxygen limitation, which often is the limiting factor in biological oxidation by attached-growth systems.

Wastewater treatment efficiency in terms of carbonaceous oxidation and nitrification can be significantly increased by the multiple staging of rotating biological contactors. A complete rotating biological contactors' system could consist of two or more parallel trains with each train consisting of multiple stages in series (1–11), Primary clarifiers are optional whereas secondary clarifiers are required for solids separation.

RBC systems can also be used for biological denitrification. Its applicability is discussed in Section 7.

2. FACTORS AFFECTING PERFORMANCE AND DESIGN

2.1. Microorganisms and Environmental Factors

The most important factor affecting performance of the rotating biological contactors is the biological slime of those microorganisms that grow on a series of thin media, such as discs, mounted side by side on a shaft. When the process is first started, the microorganisms in the wastewater begin to stick to the medium surfaces and grow there until all the media are covered with a 1/l6- to 1/8-in layer of biological slime. The attached biomass is similar to the biofilm in a trickling filter, except that the microorganisms are passed through the wastewater rather than the wastewater being passed over the microorganisms. As with all biological units, alkalinity, pH, nutrients, temperature, oxygen, biomass population balance, concentrations of pollutants, and so on must be acceptable for efficient operation.

Most organisms cannot tolerate pH levels above 9.5 or below 4.0. In general, the optimum pH for biological growth lies between 6.5 and 7.8 for carbonaceous oxidation, between 8.2 and 8.6 for nitrification (12), and between 7.2 and 7.8 for denitrification (13). An alkalinity deficit can result from nitrification; thus a supplemental alkalinity source may be required. The inorganic nutrients normally present in domestic wastewater are sufficient to assure maximum biological growth, provided that all other environmental conditions are optimum. The nutrient needs should be checked when there is a significant industrial waste contribution. A suggested ratio of BOD_5:N:P is 100:5:1. Wastewater temperatures between 13 and 32°C have no significant effect on process performance. The treatment efficiency, however, decreases with decreasing wastewater temperature below 13°C. For year-round operation in warm climates, a simple sun roof is sufficient protection; whereas for year-round operation in cold climates, rotating biological contactor plants should be weatherproofed.

To achieve high treatment efficiency, the wastewater should be maintained under aerobic conditions throughout the entire treatment system for carbonaceous oxidation and nitrification. It is suggested that a minimum of 1 to 2 mg/L of dissolved residual oxygen be maintained in the tank to prevent oxygen deficiencies from limiting the substrate removal rate (13).

Each shaft of medium operates as a completely mixed, fixed-film reactor, in which the biological growth rate and the excess biomass stripping rate are at a dynamic equilibrium. As the treated wastewater and the stripped biomass pass from stage to stage, the wastewater undergoes a progressively increased degree of treatment by the specific biomass found in each stage, which in turn adapt to the changing wastewater. Microorganisms in the initial stages of a medium, which receive the highest concentration of organic wastes, are mainly ordinary bacteria responsible for carbonaceous oxidation. Higher life forms, such as nitrifying bacteria, protozoa, rotifers, and other predators, begin to appear in subsequent stages, where the concentration of organic substances gradually decreases from stage to stage.

2.2. Media Selection and Arrangement

Common media are manufactured in the form of discs that have a specific unit surface area of 20 to 25 ft^2/ft^3 (note: 1 $ft^2/ft^3 = 3.28\ m^2/m^3$) and in the form of lattice structure that has a

specific unit surface area of 30 to 35 ft^2/ft^3. The disc-type media and the lattice-type media are generally made of styrofoam and polyethylene, respectively (14).

Standard media are available in 100,000 ft^2 shafts; high density media are available in 150,000 ft^2 shafts (4).

Basically using media with high specific unit surface area will increase the treatment efficiency. The lattice media generally perform better than the disc media because of the comparatively greater surface areas per unit volume of the former. However, the use of high specific surface area media in the first stages of treatment may result in clogging owing to the smaller clearances and, if used, requires a very low organic loading offsetting the advantage of using high density media.

The arrangement of media in a series of stages has been shown to increase treatment efficiency significantly. Current practice is to construct rotating contactor plants with at least four stages per train (15), and at least two parallel trains per plant.

Rotational velocity of the media is also important. When all stages of media in the plant rotate at the same velocity, the optimum peripheral velocity for carbonaceous oxidation and nitrification of domestic wastewater is about 0.3 m/s (60 ft/min).

It is recommended that about 40% of the media be submerged in the tank.

2.3. *Loadings and Hydraulic Parameters*

Design organic loading rates for the rotating biological reactor used for carbonaceous oxidation range from 0.5 to 1.0 kg $BOD_5/m^3/d$ (31 to 62 lb $BOD_5/1000\,ft^3/d$), and for nitrification range from 0.16 to 0.24 kg $BOD_5/m^3/d$ (10 to 15 lb $BOD_5/1000\,ft^3/d$) (4, 14, 15). The organic loading in the first stage cells should be reduced to prevent oxygen from becoming limiting, by placing more first stage cells in parallels, or by step feeding the influent wastewater to the first two cells. Use of the air driven system can certainly reduce the effect of oxygen limitation. The air drive not only moves the rotating media, but adds oxygen as well. Because the performance depends on the amount of biomass on the media, a design loading not based on media surface area is a better approach. The maximum soluble organic loading limitations on the first stage recommended by the RBC manufactures are 0.0244 kg/m^2/d (5 lb/1000 ft^2/d) (note: mass soluble BOD_5 per unit system) using standard density media. The overall soluble BOD_5 loading rates vary from 0.0049 to 0.0146 kg/m^2/d (1 to 3 lb/1000 ft^2/d). For nitrification, the recommended loading rates are from 0.0012 to 0.0024 kg/m^2/d (0.25 to 0.5 lb/1000 ft^2/d) depending on the effluent ammonia nitrogen concentration requirements. Also the stages where nitrification takes place should not have a BOD_5 concentration higher than 30 mg/L (15 mg/L soluble BOD_5).

Design hydraulic loading rates in the range of 0.03 to 0.16 $m^3/d/m^2$ (0.75 to 4.0 gpd/ft^2) of media surface area are used for secondary treatment, and in the range of 0.01 to 0.08 $m^3/d/m^2$ (0.3 to 2.0 gpd/ft^2) of media surface area are used for treatment with nitrification.

At a given hydraulic loading the wastewater will have a given retention time depending on the void fraction of the media and the size of holding tank. There is an optimum tank volume that maximizes the treatment capacity of the growth-covered surface. It has been reported (2) that the optimum tank volume for treating domestic wastewater is about 4.88 L/m^2 of media

surface (0.12 gal/ft^2 of media area), based on which the required detection time will be about 40 to 90 minutes for carbonaceous oxidation and 90 to 230 minutes for nitrification.

Although the RBC effluent contains only 50 to 150 mg/L of suspended solids, a solid-water separation facility is generally needed to follow the RBC reactor. When secondary clarifiers are provided, the recommended overflow rate at average wastewater flow is 33 m^3/m^2/d (800 gpd/ft^2) if the desired effluent suspended solids concentration ranges from 20 to 30 mg/L, and is in the range of 16 to 25 m^3/m^2/d (400 to 600 gpd/ft^2) if the desired effluent suspended solids concentration is 10 mg/L. For nitrified effluents only chemical flocculation and multiple-media filtration at 122 L/min/m^2 (3 gpm/ft^2) can assure an effluent of 5.0 mg/L suspended solids or below (16).

3. PERFORMANCE MODELS AND DESIGN PROCEDURES

3.1. US Environmental Protection Agency Model

Theoretically all performance models of trickling filters (6, 7) except the NRC model can be used for design of RBCs, provided that the constants and coefficients in the design models can be determined.

The US EPA model, Equation (1), can be applied to the RBC system (11):

$$L_e/L_o = \exp[-K_p(V/695Q)^{0.5}] \tag{1}$$

where V = media volume, ft^3; Q = wastewater design flow excluding recycle flow, MGD; L_e = reactor effluent BOD_5, mg/L; L_o = reactor influent BOD_5, mg/L; and K_p = performance measurement parameter. An average K_p value of 0.30 at 20°C has been used for many European RBC systems using primarily discs.

3.2. Modified US Environmental Protection Agency Model

Another similar formula is also frequently used for design of RBC Systems:

$$L_e/L_o = \exp[-K_t A_c/695Q)^{0.5}] \tag{2}$$

where K_t = treatability function related to surface area; and A_c = media surface area, ft^2. An average K_t value of 0.066 at 20°C has been used by many professional engineers for RBC design.

The following are the boundary conditions and design equations for RBC systems in evaluation of coefficients K_p and K_t. At wastewater temperatures above 13°C:

$$(K_p)_T = (K_p)_{20} \tag{3}$$

$$(K_t)_T = (K_t)_{20} \tag{4}$$

At wastewater temperatures below 13°C:

$$(K_p)_T = (K_p)_{20}(1.018)^{T-20} \tag{5}$$

$$(K_t)_T = (K_t)_{20}(1.018)^{T-20} \tag{6}$$

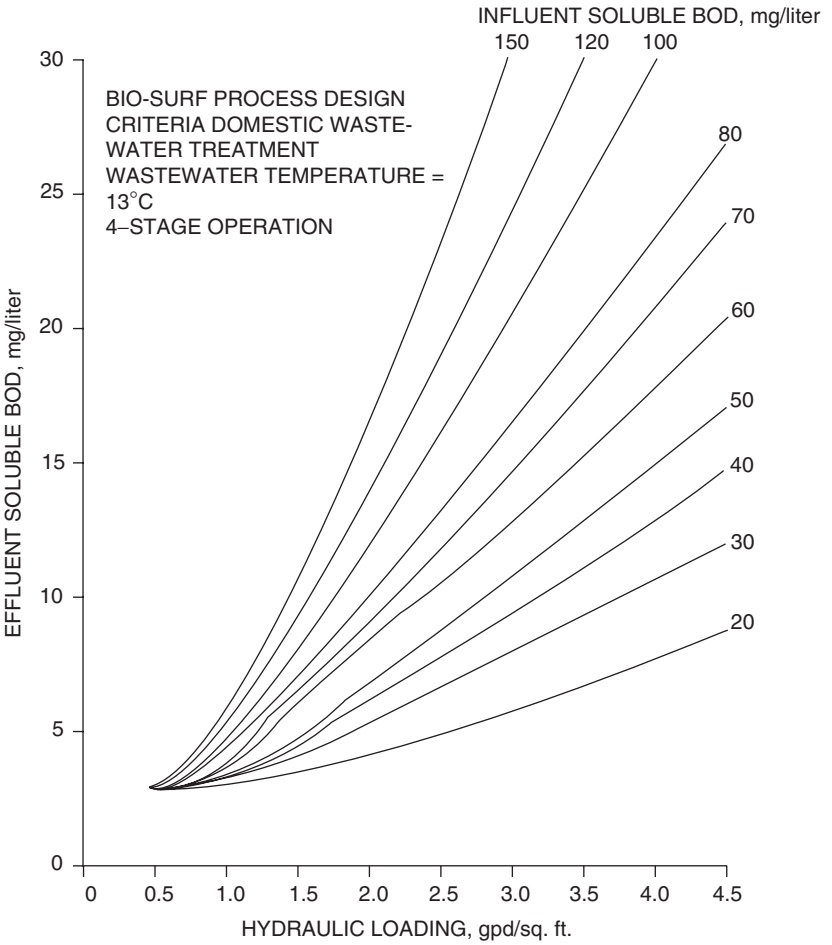

Fig. 10.2. Rotating biological media for secondary treatment (*Source*: US EPA).

where $(K_p)_T$ and $(K_p)_{20}$ are the performance measurement parameter at wastewater temperature T°C and 20°C, respectively; and $(K_t)_T$ and $(K_t)_{20}$ are the treatability parameter at wastewater temperature T°C and 20°C, respectively.

3.3. Manufacturer's Design Procedures

A design approach proposed by one of the RBC manufacturers is based on the soluble BOD_5, medium surface area, and the percent BOD_5 and/or ammonia removal efficiency. Figure 10.2 shows the design curves established by Bio-systems Division, Autotrol Corporation, for its BIO-SURF rotating disc process treating domestic wastewater at a wastewater temperature of 13°C. Temperature has no significant effect on carbonaceous oxidation by RBC at temperatures above 13°C. Below 13°C (55°F), the relationship shown in Table 10.1 would be valid.

Table 10.1
Correction Factors for RBC (bio-surf process) Hydraulic loadings at low wastewater temperatures[a]

	Wastewater temperature				
BOD_5 removal %	55°F	50°F	45°F	40°F	35°F
95	1.0	0.9	0.8	0.7	0.6
90	1.0	0.7	0.6	0.5	0.4
85	1.0	0.6	0.4	0.3	0.2

[a] Multiply hydraulic loading by the appropriate factor based on wastewater temperature and the desired treatment efficiency.

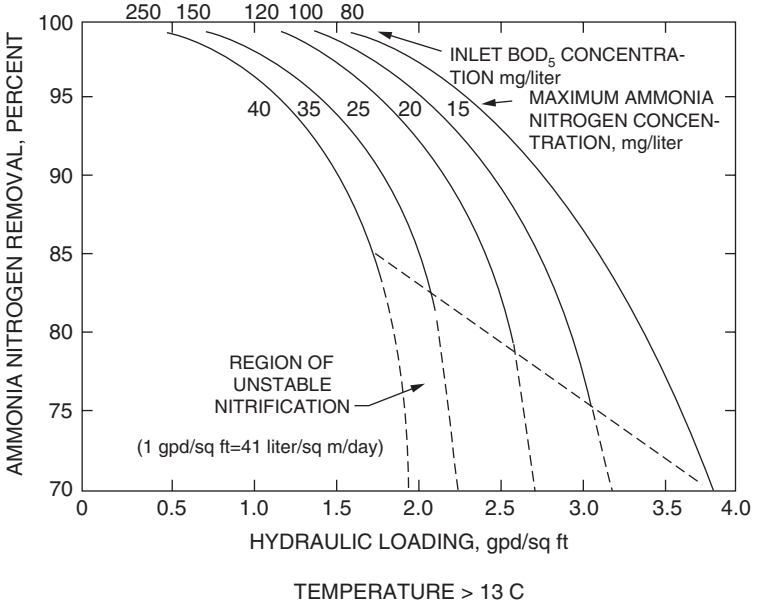

Fig. 10.3. Effect of BOD_5 concentration and hydraulic load on nitrification in RBCs (*Source*: US EPA).

It should be emphasized that the design approach whereby the organic loading is limited to the soluble organics is very liberal because of the general lack of data for this parameter. Much of the existing wastewater characteristic data indicate the effluent BOD_5 from the final clarifier will be 50% suspended material and 50% soluble. However, the influent soluble BOD_5 portion ranges from 30% to 75%.

Figure 10.3 shows the effect of BOD_5 concentration and hydraulic loading on nitrification in the RBC process at wastewater temperatures of 13°C or above. The wastewater temperature relationship for nitrification with RBC is shown in Figure 10.4, which can be used for temperature correction.

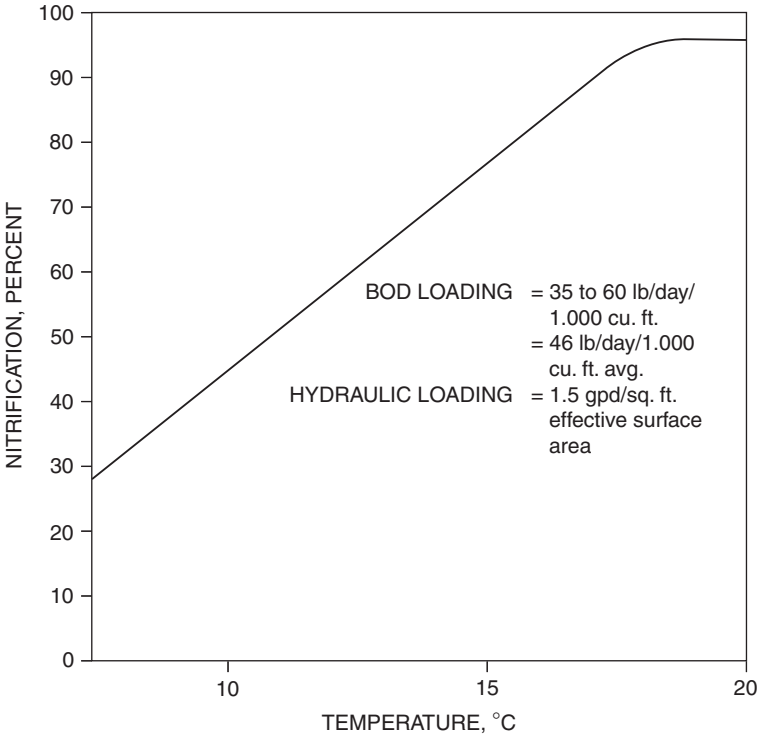

Fig. 10.4. Temperature relationship for nitrification using RBCs (*Source*: US EPA).

4. PROCESS CONTROL CONSIDERATIONS

No sludge or effluent recirculation is practiced, so there is no need for decisions on recycle rates. Suspended solids concentration of RBC effluent ranges from 50 to 150 mg/L; therefore, the amount of solids loading on the settling basin is quite low. There is no significant sludge blanket in the secondary clarifier. The settled sludge thickens to about 3% solids. If sludge is drawn off intermittently in small volumes, the 3% concentration can be maintained. If it is drawn off frequently at high rates, it will be diluted to 1% to 2% solids. The secondary sludge, however, can be recycled to the primary settling tank being thickened with primary sludge to 4% to 5%.

Where multiple units are used in each stage, loading should be kept uniform. When actual hydraulic loadings are much less than design loadings some RBC units should be idled and not filled with wastewater.

Dissolved oxygen (DO) concentration levels are maintained at 0.5 to 1 mg/L in the first stage, and at 1 to 3 mg/L in the last stage for carbonaceous oxidation. DO concentration often range from 1 to 3 mg/L in the first stage to 4 to 8 mg/L in the last stage for nitrification (4, 14).

Table 10.2 is a troubleshooting guide for RBC systems (11).

Table 10.2
Troubleshooting guide

Indicators/observations	Probable cause	Check or monitor	Solutions
1. Decreased treatment efficiency	1a. Organic overload	1a. Check peak organic loads-if less than twice the daily average, should not be the cause	1a. Improve pretreatment or expand plant
	1b. Hydraulic overload	1b. Check peak hydraulic loads—if less than trice the daily average, should not be the cause	1b. Flow equalization: eliminate source of excessive flow: balance flows among reactors
	1c. pH too high or too low	1c. Desired range is 6.5 to 8.5 for secondary treatment; 8 to 8.5 for nitrification	1c. Eliminate source of undesirable pH or add acid or base to adjust pH. When nitrifying at seven times the influent NH_3 concentrations
	1d. Low wastewater temperatures	1d. Temperatures less than 55°F will reduce efficiency	1d. If available, place added treatment units in service
2. Excessive sloughing of biomass from discs	2a. Toxic materials in influent	2a. Determine material and its source	2a. Eliminate toxic material if possible, if not, use flow equalization to reduce variations in concentration so biomass can acclimate
	2b. Excessive pH variations	2b. pH below 5 or above 10 can cause sloughing	2b. Eliminate source of pH variations or maintain control of influent pH

(*Continued*)

Table 10.2 (*Continued*)

Indicators/observations	Probable cause	Check or monitor	Solutions
3. Development of white biomass over most of disc area	3a. Septic influent or high H_2S concentrations[2]	3a. Influent odor	3a. Preareate wastewater or add sodium nitrate or hydrogen peroxide
	3b. First stage is overloaded organically	3b. Organic loading on first stage	3b. Adjust baffles between first and second stages to increase fraction of total surface area in first stage
4. Solids accumulating in reactors	4a. Inadequate pretreatment	4a. Determine if solids are grit or organic	4a. Remove solids from reactors and provide improved grit removal or primary settling
5. Shaft bearings running hot or failing	5a. Inadequate maintenance	5a. Maintenance schedules and practices	5a. Lubricate bearings per manufacturers instructions
6. Motors running hot	6a. Inadequate maintenance	6a. Oil level in speed reducer and chain drive	6a. Lubricate per manufacturers instructions
	6b. Chain drive alignment improper	6b. Alignment	6b. Align properly

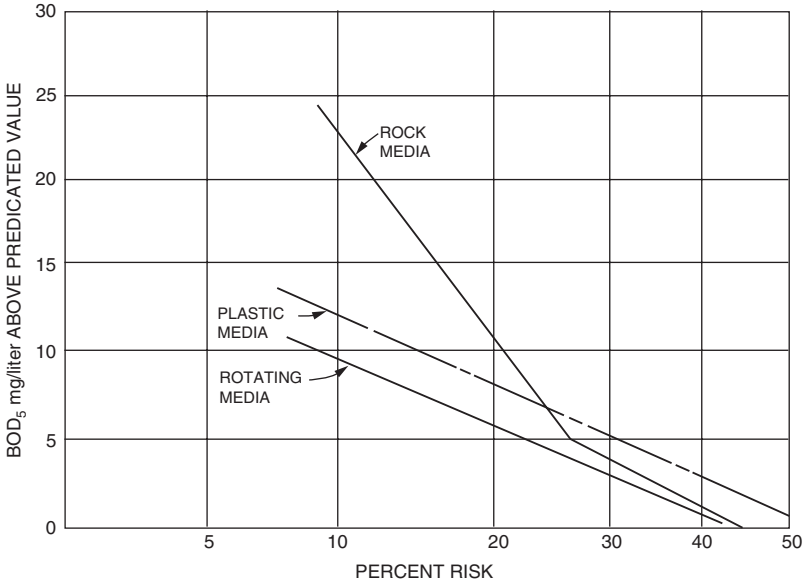

Fig. 10.5. Design risk for attached growth treatment systems (*Source*: US EPA).

5. APPLICATION, PERFORMANCE AND RELIABILITY

RBC systems are used for treating domestic and compatible industrial wastewater amenable to aerobic biological oxidation. The RBC characteristics of modular construction, low hydraulic head loss and shallow excavation allow it to be used as easily with new treatment units as with existing treatment facilities in a variety of configurations to upgrade the existing level of wastewater treatment.

The process can be used for carbonaceous oxidation and nitrification. Phosphate removal by the RBC process is similar to that achieved by other biological processes. The performance of a typical four stage RBC system with primary and secondary clarifiers is (4, 14, 15):

BOD_5 removal	80% to 90%
SS removal	80% to 90%
Phosphorus removal	0% to 30%
NH_4-N removal	Up to 95%

A degree of risk exists in meeting a specific design effluent quality when an attached-growth system is to be designed. Nevertheless, the RBC reactor is the most reliable system among all attached-growth reactors. Figure 10.5 shows the percent design risk for three common attached growth treatment media: rock media, plastic media, and rotating media. The percent risk is the probability of exceeding the design effluent concentration. To reduce the risk, the designer should select a lower design effluent BOD_5 from which to calculate the media volume.

6. LIMITATIONS AND ENVIRONMENTAL IMPACT

The process is vulnerable to climatic changes if it is not housed or covered. Treatment efficiency may be reduced significantly at wastewater temperatures below 13°C. Although an enclosure may help maintain adequate temperature, it can result in considerable wintertime condensation if heat is not then added. Supplemental aeration may be required when organic loadings to the first-stage RBC reactors are high. A supplemental alkalinity source may be required when the RBC system is used for nitrification. Use of dense media in early stages can result in media clogging.

Secondary sludge production by the RBC process is 0.4 to 0.5 kg/kg BOD_5 removed, 3000 to 4000 L wet sludge/10^6 L of wastewater, or 0.06 to 0.08 kg dry solids/m^3 of wastewater. Sludge contains about 80% volatile solids (17, 18).

7. RECENT DEVELOPMENTS IN RBC

7.1. Biodegradation of Hydrocarbon

Many types of proprietary RBC systems have been continuously developed. The advantage of RBC include (15):

1. Relative low energy consumption.
2. Simple operation and maintenance.
3. Successive treatment of the influent contaminants.

As an alternative approach to treat hydrocarbons in bioreactors, the RBC seems to be a good choice. Aerobic treatment of toluene, a typical aromatic hydrocarbon, was assessed by using a modified rotating biological contactor (19). The RBC consisted of 72 parallel discs rotating in a reservoir and was arranged in three stages, i.e. 24 discs oriented in each stage. Toluene wastewater inoculated with an enriched culture from petrochemical wastewater was fed to the RBC system. The initial toluene concentration effect on toluene removal showed a zero order mechanism.

A mixed culture of bacteria consisting of nitrifiers, heterotrophs and *Thiosphaera pantotropha* could be acclimated to achieve 99.89% removal of trichloroethylene (TCE) in wastewater at TCE loading of 0.0039 $m^3/m^2/d$ and HRT of 3.5 d (20). Carbon to nitrogen ratio of 100:20 was optimum. The system could withstand TCE shock loadings up to 0.3 $m^3/m^2/d$.

7.2. Bioremediation of Heavy Metals

Immobilized microorganisms provide a potential system for the treatment of metal contaminated waters (21). Wastewater contaminated with cadmium, copper and zinc was treated in multiple sorption-desorption cycles. Each sorption cycle extended over a period of 12 weeks at an HRT of 24 hours to determine the efficiency of the system over a protracted period of time. The removal pattern observed in the initial cycle, namely Cu $\gg$ Zn $>$ Cd, was repeated in both subsequent cycles. After completion of each cycle metals were successfully desorbed by means of an acid wash. The sorption ability of the biofilm was not adversely affected by the desorption process as evidenced by the similar metal removal rates obtained in each of

the three sorption cycles. These results suggest that RBCs can be used successfully in the treatment of high strength metal-contaminated wastewater.

A bacterial consortium capable of using metal cyanides as a source of nitrogen was used to develop a microbiological process for the detoxification of metal cyanides (viz. copper cyanide and zinc cyanide) from electroplating wastewater (22). Optimal conditions for the biodegradation of both metal-cyanide compounds were pH 7.5, temperature 35°C, inoculum size 10^9 cells/mL and glucose or sugarcane molasses requirement of 5 mM or 0.6 mL/L. Metal precipitates obtained during metal-cyanide biodegradation were identified as metal-hydroxides. When the treatment was carried out in a 27 L rotating biological contactor in continuous mode, the system was able to achieve >99.9% removal of 0.5 mM metal cyanide (ca. 52 mg/L cyanide and 30 to 40 mg/L copper/zinc) in 15 hours with sugarcane molasses as carbon source. The RBC treated effluent was found to be safe for discharge in the environment as confirmed by chemical analysis and fish bioassay studies.

7.3. Denitrification

Several studies have employed RBC systems for nitrogen removal (23, 24). A pilot rotating biological contactor was used to remove nitrate-nitrogen from groundwater using methanol, ethanol and acetic acid as carbon sources. The addition of organic compounds enhanced nitrogen removal, and acetic acid was shown to be the best additive (25, 26). The reactor achieved a nitrate removal efficiency of 99 and 83% at loading rates of 76 and 490 $mg/m^2/h$ respectively with a flow rate of 2.5 L/min at20°C.

Teixeira et al. (27) used completely submersed discs for denitrification. Two RBCs, one with completely submersed discs (100% submergence) and the other with partially submersed discs (64.5%), were operated under the same conditions. Their performance was evaluated in terms of denitrification efficiency as well as biofilm characteristics, composition and activity. As far as the denitrification process is concerned, the RBC with a completely submersed biofilm was more efficient than the other but had a longer delay in start-up. The biofilm of both reactors was very thick (>0.6 mm) but with different structures. Biofilm activity seems to be directly dependent on the biofilm structure, namely on the degree of hydration.

7.4. Improvement of RBC Design

The principle of the RBC originated in the early part of the last century and today there are many thousands of units operating worldwide. However, the RBC has been often plagued with mechanical deficiencies since its conception. Mba et al. (28) presented a brief insight into some of the main mechanical defects associated with RBCs. Some reasons for mechanical failures are attributed to poor engineering design, low frequency corrosion fatigue and microbiologically influenced corrosion. Having the benefit of a thorough understanding of the mechanisms and reasons for mechanical failure, an improvement in designing RBCs was proposed resulting in a new generation of RBC designs for long life operation (29). The choice of material for the RBC is an important design feature. A programmed study of biofilms growth in rotating biological reactors that consisted of discs made from different materials showed that for activated carbon, sand, glass particles and diatomaceous earth, carbon coated

discs were the best for the biofilm growth (30). In a recycle facility the relatively poor effluent from the RBC was used as a carbon source allowing the possibility of denitrification though constructed reed beds. This operation offers the possibility of achieving high quality effluents (31). The use of a jet mixed separator, which has a series of porous plates inserted in the channel perpendicular to the flow, resulted in enhanced flocculation and sedimentation. According to Watanabe et al. (32) who explained the process in detail, the water passes through holes in the plates, creating jets, which in turn mix the water resulting in simultaneous flocculation of suspended particles and their subsequent settling and removal.

7.5. Domestic Wastewater Treatment and Purification

Boumansour and Vasel (33) developed a new tracer gas method using propane to measure oxygen transfer and enhancement factor on RBC. Thus, the method can be used to determine a true enhancement factor without using reference clean discs. The enhancement factor has been successfully correlated to the oxygen consumption kinetics, as one might expect from theoretical considerations.

Serial application of the RBC system has been demonstrated in aerobic treatment processes such as decolorization, nitrification and pathogenic bacteria removal from domestic wastewater. The combination of anaerobic-aerobic-anoxic systems for the removal of nitrogen and phosphorus from domestic wastewater has been successfully demonstrated (34, 35). The careful characterization of ciliate communities inhabiting RBC biofilms improved the knowledge of the role of microorganisms in wastewater treatment (36). Mathematical models were developed that help in predicting RBC effluent qualities (37). The use of polyurethane foam to support the biofilm of immobilized cells on their surface increased the RBC performance (38).

Studies have been focused on simultaneous nutrient removal in RBC systems (39, 40). *Thiosphaera pantotropha* has been shown to be capable of simultaneous heterotrophic nitrification and aerobic denitrification thereby helping the steps of carbon oxidation, nitrification and denitrification to be carried out concurrently exhibiting high simultaneous removal of carbon and nitrogen in fully aerobic conditions (41). Purification of RBC treated domestic wastewater for reuse (42) has been demonstrated to produce effluents with total nitrogen less than 10 mg/L, which meets the regulation criterion for nitrogen (43–44).

8. DESIGN EXAMPLES

8.1. Example 1

A rotating biological contactor (RBC) system is to be designed for wastewater treatment. The desired effluent BOD_5 concentration is 30 mg/L. If the design risk is chosen to be 20%, determine the effluent BOD_5 concentration to be actually used in design.

Solution

The risk analysis shown in Figure 10.5 indicates that the BOD_5 concentration will be 6 mg/L above the predicted value at 20% risk when the rotating media are to be used. The

effluent BOD, concentration to be actually used in design is equal to:

$$30\,\text{mg/L} - 6\,\text{mg/L} = 24\,\text{mg/L}$$

8.2. *Example 2*

A three-stage RBC system is to be designed based on the following environmental and design conditions:

(a) RBC influent BOD_5 concentration (L_0) = 135 mg/L
(b) Desired RBC effluent BOD_5 concentration = 30 mg/L
(c) Design risk = 20%
(d) Wastewater influent flow (average daily) = 1 MGD
(e) Wastewater temperature (T) = 10°C
(f) Design model, use Equation (52)
(g) Treatability parameter (K_t) = 0.066 at 20°C
(h) RBC system = three equal-size stages; the third stage is designed with media having 50% more surface area than the first two stages
(i) Design flow (Q) = peak month flow 1.45 (average day flow)

Determine the following:

(a) Design effluent concentration (L_e)
(b) Treatability parameter $(K_t)_T$
(c) Total surface area (A_c) of the rotating media, ft^2
(d) Surface area of each stage, ft^2
(e) Hydraulic loading (q) under average day flow conditions
(f) Hydraulic loading (q) under peak month flow conditions
(g) Organic loading on the first stage under peak month flow conditions
(h) Organic loading of the entire system under average day flow conditions

Solution

(a) Design effluent BOD_5 concentration has been determined in Example 1: $L_e = 24\,\text{mg/L}$

(b)
$$(K_t)_T = (K_t)_{10} = (K_t)_{20} \times (1.018)^{10-20} \quad (6)$$
$$= 0.066 \times 1.018^{-10}$$
$$= 0.055$$

(c)
$$L_e/L_o = \exp[-K_t(A_c/695Q)^{0.5}] \quad (2)$$
$$24/135 = \exp[-0.055(A_c/695 \times 1.45 \times 1)^{0.5}]$$
$$A_c = [\ln(L_e/L_o)K_t]^2(695Q) \quad (7)$$
$$= [\ln(24/135)/0.055]^2(695 \times 1.45 \times 1)$$
$$= 993{,}855\,ft^2$$

(d) The distribution of media surface area among the three equal size stages would be as follows:

Stage	Furnished surface area, ft^2
1st	300,000
2nd	300,000
3rd	450,000
Total	1,050,000 > 993.855

(e) Hydraulic loading (q) under average day flow conditions $= 1.00 \times 10^6$ gpd/1,050,000 ft^2 = 0.95 gpd/ft^2

(f) Hydraulic loading (q) under peak month flow conditions $= 1.45 \times 10^6$ gpd/1,050,000 ft^2 = 1.38 gpd/ft^2

(g) Organic loading on the first stage under peak month conditions

$$= (1.35 \times 8.34)(1.45)/300 \text{ lb } BOD_5/1000\,ft^2/d = 5.4 \text{ lb } BOD_5/1000\,ft^2/d$$

which is very close to, but slightly higher than 5 lb BOD_5/1000 ft^2/d, which is the manufacturer's recommended maximum organic loading for the first stage of an air-driven systems (see Section 2.3). In light of this it would be appropriate to provide operational flexibility to feed some of the influent wastewater to the second stage.

(h) BOD_5 loading of overall system at average day flow $= (135 \times 8.34)(1)/1{,}050 =$ 1.07 lb/1000 ft^2/d

8.3. Example 3

Solve Example 2 again using the US EPA model Equation (1), taking the performance measurement parameter (K_p) to be equal to 0.30 at 20°C.

Solution

(a) $L_e = 24$ mg/L

(b)
$$(K_p)_T = (K_p)_{10} = (K_p)_{20} \times (1.018)^{10-20} \tag{5}$$
$$= 0.30 \times 1.018^{-10}$$
$$= 0.25$$

(c)
$$L_e/L_o = \exp[-Kp(V/695Q)^{0.5}] \tag{1}$$
$$24/135 = \exp[-0.25(V/695 \times 1.45 \times 1)^{0.5}]$$
$$V = [\ln(L_e/L_o)/K_p]^2(695Q) \tag{8}$$
$$= [\ln(24/135/0.25)^2(695 \times 1.45 \times 1)$$
$$= 48{,}103\,ft^3$$

According to Section 2.3, common disc media have a specific surface area of 20–25 ft^2/ft^3. Suppose the disc media having 20.8 ft^2/ft^3 is chosen; the total surface area of the media (A_c) is

Table 10.3
Design of RBC for secondary treatment at various wastewater flows[a]

Average day flow, MGD	1.0	5.0	10.0	50.0	100.0
Peak month flow, MGD	1.45	7.1	13.8	58.0	116.0
Peak day flow. MGD	3.5	13.5	25.0	100.0	200.0
Media surface area, $10^6\,ft^2$					
First stage	0.30	1.30	2.50	10.60	21.20
Second stage	0.30	1.30	2.50	10.60	21.20
Third stage	0.45	2.10	3.90	15.90	31.80
Total	1.05	4.70	8.90	37.10	74.20
Hydraulic loading, gpd/ft^2					
Average day flow	0.95	1.06	1.12	1.35	1.35
Peak month flow	1.38	1.51	1.55	1.56	1.56
BOD loading, $lb/1000\,ft^2/d$					
First stage @ peak month	5.40	6.10	6.21	6.15	6.15
Overall @ average day	1.07	1.20	1.26	1.52	1.52

[a] RBC influent BOD_5 concentration = 135 mg/L. Wastewater temperature = 10°C.

equal to 48,103 × 20.79 or 1,000,000 ft^2, which is very close to that determined in Example 2. The rest of the calculations are identical to those of steps (d), (e), (f), and (g) in the solution of Example 2.

8.4. Example 4

Design an RBC system for secondary treatment at the following wastewater flows:

Average day flow, MGD	5.0	10.0	50.0	100.0
Peak month flow, MGD	7.1	13.8	58.0	116.0
Peak day flow, MGD	3.5	25.0	100.0	200.0

assuming all other environmental and design conditions are identical to that of Example 2.

Solution

The solutions of Examples 2 and 4 are summarized in Table 10.3. It should be noted that the organic loadings on the first stage under peak month flow conditions are all over 5 lb $BOD_5/1000\,ft^2/d$; therefore, a design should provide for flexibility in distributing influent wastewater to first two stages of media.

8.5. Example 5

An RBC system is to be designed for both carbonaceous oxidation and nitrification. The stages of the RBC wherein nitrification occurs are considered to begin when the BOD_5 loading

falls to 15 lb/d/1000 ft^3 of the remaining volume. From the data presented for nitrification using the RBC, a hydraulic loading of less than 1.25 gpd/ft^2 and an organic loading below 15 lb/1000 ft^3 are reasonable values for design of second stage units (see Section 2.3.). Additional environmental and design conditions are assumed:

(a) Wastewater average day flow = 1 MGD
(b) Design flow (Q) = peak month flow = 1.45 (average day flow)
(c) RBC influent BOD_5 concentration = 135 mg/L
(d) Rotating media used = lattice media
(e) Specific surface area of media = 32 ft^2/ft^3
(f) Design model, use Equation (2)
(g) RBC system = 4 stages
(h) Wastewater temperature = 10°C

Determine the following:

(a) BOD_5 concentration of partially treated wastewater at which nitrification begins.
(b) Surface area of the lattice media for carbonaceous oxidation before nitrification begins.
(c) Surface area of the lattice media for nitrification.
(d) Arrangement of the staging of the media.
(e) Hydraulic loading at peak month flow.
(f) Hydraulic loading at average day flow.
(g) Organic loading on the first stage under peak month conditions.

Solution

(a) Design wastewater flow (Q) = 1 × 1.45 = 1.45 MGD

At a hydraulic loading rate (q) of 1.25 gpd/ft^2:
Volume of media (V) = $1.45 \times 10^6/(1.25 \times 32) = 36{,}300$ ft^3
At an organic loading rate of 15 lb/1000 ft^3/d:
BOD_5 concentration at which nitrification begins

$$= (15/1000)(36{,}300)/(8.34 \times 1.45)$$

$$= 45 \text{ mg/L} \ldots L_e \text{ for carbonaceous oxidation}$$

(b) K_t at 20°C = 0.066 (Section 3)
K_t at 10°C = 0.055 (Example 2)

$$A_c = [\ln(L_e/L_0)/K_t]^2(695Q) \tag{57}$$

$$= [\ln(45/135)/0.055]^2(695 \times 1.45)$$

$$= 402{,}084 \text{ ft}^2 \text{for carbonaceous oxidation}$$

(c) $A_c = Q/q = (1.45 \times 10^6 \text{ gpd})(1.25 \text{ gpd/ft}^2) = 1{,}160{,}000$ ft^2 for nitrification

(d) Total rotating media provided

$$= 402{,}084 + 1{,}160{,}000 = 1{,}562.084 \text{ ft}^2$$

Possible staging of the media is as follows:

Stage	Furnished Surface Area, ft^2
1st	340,000
2nd	340,000
3rd	450,000
4th	450,000
Total	1,580,000 > 1,562,084

(e) Actual hydraulic loading at peak month flow:

$$q = (1.45 \times 10^6 \text{ gpd})/1{,}580{,}000 \text{ ft}^2 = 0.92 \text{ gpd/ft}^2$$

(f) Actual hydraulic loading at average day flow

$$q = (10^6 \text{ gpd})/1{,}580{,}000 \text{ ft}^2 = 0.63 \text{ gpd/ft}^2$$

(g) Organic loading on the first stage under peak month conditions

$$= (135 \times 8.34 \times 1.45 \text{ lb/d})/340(1000 \text{ ft}^2)$$
$$= 4.8 \text{ lb } BOD_5/1000 \text{ ft}^2/\text{d}$$

which is less than the manufacturer's suggested loading, 5 lb $BOD_5/1000$ ft^2/d.
Organic loading on the first stage under average day conditions

$$= (135 \times 8.34 \times 1 \text{ lb/d})/340(1000 \text{ ft}^2)$$
$$= 3.3 \text{ lb } BOD_5/1000 \text{ ft}^2/\text{d}$$

8.6. Example 6

Design the RBC systems for both carbonaceous oxidation and nitrification at the following wastewater flows:

Average day flow, MGD	5.0	10.0	50.0	100.0
Peak month flow, MGD	7.1	13.8	58.0	116.0
Peak day flow, MGD	13.5	25.0	100.0	200.0

assuming all other environmental and design conditions are identical to that of Example 5.

Solution

The solutions of both Example 5 and Example 6 are summarized in Table 10.4.

The organic loading on the first stage under peak month flow conditions are very close to 5 lb $BOD_5/1000$ ft^2/d, which is recommended by the manufacturer.

8.7. Example 7

Calculate the rotating-disc surface area required for secondary treatment of 1 MGD of raw domestic wastewater having an ultimate BOD_u 230.0 mg/L. The effluent BOD_5 specified is 20 mg/L when the wastewater temperature is 45°F. It is assumed that the treatment efficiency

Table 10.4
Design of RBCs for carbonaceous oxidation and nitrification at various wastewater flows[a]

Average day flow, MGD	1.0	5.0	10.0	50.0	100.0
Peak month flow, MGD	1.45	7.1	13.8	58.0	116.0
Peak day flow, MGD	3.5	13.5	25.0	100.0	200.0
Media surface area, $10^6\,ft^2$					
First stage	0.34	1.60	3.00	12.85	25.60
Second stage	0.34	1.60	3.00	12.85	25.60
Third stagc	0.45	2.30	4.44	18.40	36.90
Fourth stage	0.45	2.30	4.44	18.40	36.90
Total	1.58	7.80	14.88	62.50	125.00
Hydraulic loading, gpd/ft^2					
Average day flow	0.63	0.64	0.67	0.80	0.80
Peak month how	0.92	0.91	0.93	0.93	0.93
BOD loading, $lb/1000\,ft^2/d$					
First stage @ average month	3.30	3.52	3.75	4.38	4.40
First stage @ peak month	4.80	5.0	5.18	5.08	5.10

[a] RBC influent BOD_5 concentration = 135 mg/L. Wastewater temperature = 10°C.

of primary clarifier is 35%, and the soluble portion of the BOD_5 in clarifier effluent is 67%. Use the manufacturer's procedure for sizing.

Solution

Primary effluent $BOD_5 = 230(1 - 0.35) = 150\,\text{mg/L}$
Soluble BOD_5 of RBC influent $= 150 \times 0.67 = 100\,\text{mg/L}$
Soluble BOD_5 of RBC effluent $= 20 \times 0.50 = 10\,\text{mg/L}$
Hydraulic loading $= 1.75\,\text{gpd/ft}^2$ at 13°C (from Figure 10.2)
Desired RBC efficiency $= (100 - 10)/100 = 0.90 = 90\%$
Correction factor = 0.6 (from Table 10.1)
Corrected hydraulic loading at 45°F

$$q = 1.75 \times 0.6 = 1.05\,\text{gpd/ft}^2$$

Disc surface area required

$$A_c = 1{,}000{,}000\,\text{gpd}/(1.05\,\text{gpd/ft}^2)$$
$$= 952{,}381\,\text{ft}^2$$

8.8. Example 8

The design and operating data of an existing RBC system are listed below:

(a) 4-stage RBC system
(b) Effective surface (single-stage) $= 39{,}062\,\text{ft}^2$

(c) Influent soluble BOD_5 concentration = 100 mg/L
(d) Effluent soluble BOD_5 concentration = 20 mg/L
(e) Wastewater flow = 0.5 MGD

Determine and discuss:

(a) The hydraulic loading for the existing system.
(b) The efficiency of BOD_5 removal.

Solution

(a) Hydraulic loading

$$\begin{aligned} q &= (\text{total flow})/[(\text{number of stages})(\text{area/stage})] \\ &= (5 \times 10^5 \text{ gpd})/[(4)(39{,}062)] \\ &= 3.2 \text{ gpd/ft}^2 \end{aligned}$$

As shown in Figure 10.2, a loading rate of 3.2 gpd/ft^2 with an influent BOD_5 of 100 mg/L should produce an effluent BOD_5 around 23 mg/L.

(b) Percent BOD, removal

$$= 100(100 - 20)/100 = 80\%$$

The effluent BOD_5, 20 mg/L, is slightly better than the expected value, 23 mg/L.

If the BOD_5 removal is much less than the expected value, the troubleshooting guide indicated in Table 10.2 should be consulted.

8.9. Example 9

Discuss the applicability of RBC system for biological nitrification and denitrification.

Solution

A complete rotating biological contactor system consisting of carbonaceous oxidation, nitrification and denitrification, can be used for nitrogen conversion and removal. The denitrification can be accomplished by the operation of a RBC disc shaft in a completely submerged position with addition of a carbon source, such as methanol. Anaerobic denitrifying bacteria will develop naturally on the surfaces of the denitrification discs without the need for an intermediate clarifier between the nitrification units and the denitrification units. It should be noted that intermediate clarifier is also not needed between carbonaceous oxidation and nitrification steps.

Designing the RBC system for nitrification is the first step in designing for denitrification. The RBC nitrification unit consistently produces an effluent of about 1 to 2 mg/L ammonia nitrogen. The total nitrogen content of the nitrification unit's effluent must be determined to establish the degree of denitrification required. It has been reported (3) that a completely submerged RBC assembly can be loaded hydraulically at 1.3 gpd/ft^2 with a detention time of 5 hours for denitrification. This is only a conservative approach to denitrification design until more specific criteria are established through pilot plant testing.

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NOMENCLATURE

A_c = media surface area, L^2
K_p = performance measurement parameter
$(K_p)_T$ = performance measurement parameter at temperature T, °C
$(K_p)_{20}$ = performance measurement parameter at 20°C
K_t = treatability function related to surface area
$(K_t)_T$ = treatability parameter at temperature T, °C
$(K_t)_{20}$ = treatability parameter at 20°C
L_e = reactor effluent BOD_5, M/L^3
L_o = reactor influent BOD_5, M/L^3
q = hydraulic loading rate, $L^3/T/L^2$
Q = wastewater design flow, L^3/T
T = temperature, °C
V = media volume, L^3

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11
Sequencing Batch Reactors

Lawrence K. Wang and Yan Li

CONTENTS

Abstract A sequencing batch reactor (SBR) can be either a biological SBR (BIO-SBR) or a physicochemical SBR (PC-SBR). BIO-SBR includes traditional sedimentation biological SBR, innovative flotation biological SBR (BIO-DAF-SBR), innovative membrane biological SBR (MBR-SBR), aerobic digestion SBR (AD-SBR), etc. All PC-SBR are innovative processes including at least sedimentation PC-SBR (PC-SED-SBR), flotation PC-SBR (PC-DAF-SBR), membrane PC-SBR (PC-membrance-SBR), granular activated carbon PC-SBR (PC-GAC-SBR), powdered activated carbon PC-SBR (PC-PAC-SBR), and ion exchange PC-SBR (PC-IX-SBR). Although all BIO-SBR and PC-SBR processes are introduced, special emphasis of this chapter is placed on traditional BIO-SBR which can be used for wastewater treatment aerobically (bio-oxidation and nitrification), or anoxically/anaerobically (denitrification). The advantages, disadvantages, applications, performance, theory, operation, maintenance, design of traditional biological SBR are presented in detail. Eleven design examples and case studies for both BIO-SBR and PC-SBR are reported.

Key Words Batch suspended growth biological system • sequencing batch reactor (SBR) • biological SBR (BIO-SBR) • physicochemical SBR (PC-SBR) • flotation biological SBR (BIO-DAF-SBR) • innovative membrane biological SBR (MBR-SBR) • aerobic digestion SBR

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(AD-SBR) • sedimentation PC-SBR (PC-SED-SBR) • flotation PC-SBR (PC-DAF-SBR) • membrane PC-SBR (PC-membrance-SBR) • granular activated carbon PC-SBR (PC-GAC-SBR) • powdered activated carbon PC-SBR (PC-PAC-SBR) • and ion exchange PC-SBR (PC-IX-SBR) • theory • operation • performance • design examples • case studies.

1. HISTORICAL DEVELOPMENT AND GENERAL PROCESS DESCRIPTIONS

1.1. All Sequencing Batch Reactor Processes

Every industry periodically encounters an innovative technology that creates a widespread interest and revolutionary applications within that industry. Sequencing batch reactor (SBR) processes represent such a technology in the field of environmental engineering. Wesley M. Shubert, consultant to Aqua-Aerobic System Inc. (1), is considered to be a pioneer of the SBR. Mr. Shubert has conducted extensive research, and has completed many SBR process designs for municipal and industrial applications.

Sequencing batch reactor processes actually represent a very elementary form of treatment process systems known as fill and draw, similar to the household washing machine operation. Entire treatment is accomplished in one reactor.

Various sequencing batch reactor processes have been developed for a wide range of environmental applications including: potable/industrial waster treatment, municipal/industrial wastewater treatment, and solid waste handling and treatment.

The sequencing batch reactor processes can be biological, physicochemical or biological-physicochemical.

The process sequence usually includes the steps of equalization/filling, mixing, reaction, clarification, and decanting. The clarification step can be sedimentation, flotation, or membrane separation. The biological process step can be aerobic, anoxic or anaerobic, and can be a suspended growth system or an attached growth system.

1.2. Physicochemical SBR Process Involving Sedimentation Clarification

One of many sequencing batch processes developed by Drs. Lawrence K. Wang, Lubomyr Kurylko and Mu Hao Sung Wang in 1994 (2) is a physicochemical SBR process which can be used for potable water purification, industrial water treatment, industrial effluent treatment, and groundwater decontamination (3). The process itself is very simple and can be understood by all environmental and sanitary engineers because it is very similar to the well-known standard jar-test. The process steps include filling/equalization, chemical addition, mixing, reaction/flocculation, clarification (sedimentation) and decanting.

Although the process is new, all existing commercial sequencing batch reactor (SBR) process equipment and instrumentation can be adopted for practical applications.

1.3. Aerobic-Anoxic Biological SBR Process Involving Sedimentation Clarification

The most famous sequencing batch process is biological SBR developed and perfected by Wesley M. Shubert in 1986 (1) and Kenneth A. Mikkelson in 1995 (4). Aqua-Aerobic Systems, Inc. (5, 6) alone has more than 600 installed systems internationally. Biological SBR systems are proven to be efficient alternatives to conventional flow-through methods.

Original classic SBR is an aerobic, or an aerobic-anoxic process which has no enclosure on the top, and uses sedimentation for clarification. Basically SBR is very similar to conventional activated sludge process, except that SBR adopts a batch biological reactor, which can be applied to carbonaceous oxidation, nitrification and de-nitrification. Aqua-Aerobic System, Inc. has eight US Patents for SBR: Nos. 4695376, 4422771, 4956100, 5358644, 4997557, 5228996, 4883602 and 6019898.

This is a well-developed biological wastewater treatment process, which has been studied and accepted internationally (1, 3, 5, 7–16). Nalasco, Irvine and Monoharan presented their review, conclusions and favorable comments on SBR in 1998 (10). This chapter will introduce this classic SBR process in detail.

A contact stabilization SBR process has been studied experimentally in the laboratory by Liu (16) and piloted for municipal sewage treatment at 9 to 12°C by Pankivskyi (14) under direct supervision of Dr. Milos Krofta and Dr. Lawrence K. Wang.

Fluidyne has claimed development of an innovative Integrated Surge Anoxic Mix SBR system (15). The aerobic-anoxic system, however, was also developed by W. M. Shubert in 1986 for carbonaceous oxidation, nitrification and denitrification. (1) under a project sponsored by Aqua-Aerobic Systems, Inc.

Bio-augmentation is a process step involving the addition of microorganisms to a waste treatment process unit (such as SBR), an ex-situ hazardous material processing system, or a contaminated site to degrade specific contaminants readily and biologically. Application of bio-augmentation to biological SBR is highly recommended.

Nitroglycerin, also known as glycerol trinitrate (GTN) is manufactured for use as an explosive in double-base gun and rocket propellants and for use as a pharmaceutical vasodilator. As a result of its industrial applications, GTN is frequently formed in the waste streams, contaminated soils and groundwater near munitions manufacturing facilities and pharmaceutical facilities. It was concluded by Accashian, Smet and Kim (9) that complete removal of GTN and glycerol dinitrates (GDN) was contingent on addition of a GTN-adapted inoculum in an SBR reactor.

In case the air emission from an SBR reactor is hazardous or odorous, an enclosure for SBR and a supplemental air purification system has been developed (2).

1.4. Aerobic-Anoxic Biological DAF-SBR Process Involving Flotation Clarification

In 1994 Wang, Kurylko and Wang (2) also developed an innovative aerobic-anoxic biological dissolved air flotation SBR process (BIO-DAF-SBR process) which uses flotation for clarification.

This newly developed biological SBR process (2) is very similar to the traditional innovative biological SBR process, except that innovative dissolved air flotation (DAF) clarification is used instead of traditional sedimentation clarification for further cost saving.

Dissolved air flotation is a cost effective liquid treatment process in which pressurization is applied to dissolve air in water (under high pressure), then depressurization is applied to release and generate extremely fine air bubbles (under normal one atmospheric pressure) for separation of impurities, pollutants, light nonaqueous phase liquids (LNAPLs) and other suspended particles from a contaminated water.

The combined application of SBR and DAF may reduce clarification time from 1 to 4 hours to about 10 to 15 minutes, and in turn, may reduce capital and O and M costs.

The biological DAF-SBR process has been demonstrated for groundwater decontamination (3) and municipal sewage treatment (14). This process system may be further equipped with an enclosure and air purification means when it is used for treating explosive, hazardous and odorous wastewater (2).

Powdered activated carbon (PAC) may be dosed to the bioreactor to increase treatment efficiency (3), and alum or ferric chloride may be dosed for phosphate removal. In such a case, it is a physicochemical and biological DAF-SBR process (or PC-BIO-DAF-SBR process).

Theory and principles of aerobic and anoxic activated sludge process operation can be found elsewhere (17–26).

1.5. Physicochemical DAF-SBR Process Involving Flotation Clarification

A physicochemical DAF-SBR process (or PC-DAF-SBR) has also been developed by Wang, Kurylko and Wang (2) and was demonstrated for treatment of an electroplating effluent containing hazardous heavy metals and volatile organic compounds (VOCs).

The same physicochemical DAF-SBR process is also feasible for treating potable water, contaminated water (3), sewage and other industrial effluents. The process equipment and process steps of physicochemical DAF-SBR are very similar to that of physicochemical sedimentation SBR (PC-SED-SBR), except that the former adopts DAF clarification, and the later adopts sedimentation clarification.

This process equipment (PC-DAF-SBR) may also be further equipped with an enclosure and air purification means when it is used for treating hazardous and odorous wastewater. For portable water purification and industrial effluent treatment, addition of PAC may enhance final polishing efficiency.

1.6. Biological Membrane-Bioreactor-(MBR-SBR) Process

A new MBR-SBR process involving the combination of two unit processes (SBR and membrane bioreactor) has been attempted by Choo and Stensel in 2000 (27). Specifically, performance of a laboratory scale MBR-SBR using a microfiltration membrane for secondary effluent filtration (instead of traditional secondary sedimentation) was studied in terms of COD removal, nitrogen removal, and membrane permeability during long term continuous operation for treating a synthetic wastewater. During the entire one year period of operation, effluent turbidity was less than 0.2 NTU, and substantial removal of COD and nitrogen was achieved. Average effluent COD, ammonia nitrogen, nitrate/nitrite nitrogen were in the range of 2.6 to 8.5 mg/L, < 0.1 mg/L and 3.2 to 5.6 mg/L, respectively.

1.7. Biological Anaerobic SBR Process

Herum and Dague (28) and Schmit and Dague (29) studied biological anaerobic SBR process in the absence of oxygen. The effect of applied vacuum on the performance of the anaerobic SBR was investigated (28). Anaerobic SBR was used to treat a swine waste successfully (29).

Research is continuing at Iowa State University on this new biological process under US Patent No. 5,185,079. The researchers at the University have shown that this process is capable of achieving high COD removal rates over a wide range of temperatures when treating various agricultural and industrial waste streams. The project was initiated by Pidaparti (30) at the university.

1.8. Biofilm SBR Process

All biological SBR processes introduced in Sections 1.3, 1.4, 1.6 and 1.7 are biological suspended growth systems, in which microorganisms are suspended in biological reactors by mixing and aeration.

A new biofilm SBR process recently studied by many researchers (31–33) is of a biological attached growth system in which microorganisms attach on the surface of a medium. The biofilm SBR process is similar to conventional plastic media trickling filter process, except that the former is a batch process and the latter is a continuous process. The researchers' experimental findings are presented below.

Fitch et al. (31) studied TCE degradation by *Methylosinus trichosporium* in a biofilm SBR unit operated with separate growth and degradation phases. The apparent pseudo-first –order degradation rate constant was improved from 0.008 to 0.1 L/mg-d by using mutant *M. trichosporium* with a soluble methane monooxygenase. However, only short degradation cycles could be sustained with 100-g/L influent TCE.

Garzon-Zuniga and Gonzalez-Martinez (32) studied the integration of phosphorus and nitrogen removal in a biofilm SBR unit with four reaction phases: anaerobic/aerobic/anoxic/aerobic. Optimal operating conditions in a 1000-L reactor filled with Pall-Rings were achieved after 615 days. Removal of COD, phosphates and NH_4^+-N were $89 \pm 1\%$, $75 \pm 15\%$, and $87 \pm 10\%$, respectively.

A biofilm SBR process was operated by Munoz-Colunga and Gonzalez-Martinez (33) for 400 days to examine the effects of operating strategy on nutrient removal. Four stages (filling, anaerobic phase, aerobic phase, and draw down) were varied, and nutrients were simulated with a molasses and phosphate solution. The highest COD and PO_4-P removal rate were obtained with 12-hour cycles, 37% anaerobic and 63% aerobic.

1.9. Solid Waste SBR Digestion Process

Wang and Wang (34) developed a SBR digestion process for treatment of solid waste. Specifically Wang and Wang's invention (34) relates to a multistage, single sequencing batch reactor (SBR) digestion system for disinfecting, washing, and concentrating solid wastes which include contaminated soil, highly concentrated biosolids, spent animal bedding, low-radioactive wastes and chemically, enzymatically degradable spent diapers, and medical wastes.

In operation, a solid waste is fed into the reactor of this SBR digestion process apparatus (34). At least one disinfecting and chemical agent (including cationic surface active agent known as quaternary ammonium compound; chemical oxidation agent; and/or enzymatic digestion agent) is fed together with water (as needed) into this reactor to disinfect and digest the solid waste with the assistance of a mechanical mixing means inside the reactor.

A chemical is fed to the same reactor as needed. After the period of feeding, disinfecting, digesting, and mixing is over, the process water is drained. The disinfected and processed wet solid waste is discharged to an expulsion chamber means which removes residual water and compacts the disinfected and processed solid waste.

The drained process water and the removed residual water are discharged either into a municipal sewer system for proper disposal, or into a chemical neutralizer means for pretreatment before its sewer discharge, or for production of a fertilizer aid containing nitrogen and phosphorus.

The batch operated reactor (34) is very similar to a household washing machine. In case an enzymatic digestion agent is used, certain solid waste, such as the spent degradable diapers in the hospitals, can dissolve in water resulting in solid waste volume reduction.

1.10. Ion Exchange-SBR Process

An ion change-SBR process (IX-SBR) was developed by Wang, Kurylko and Wang in 1996 (2) for removal of heavy metals from industrial effluents in Pittsfield, MA. The process is similar to conventional biological SBR process in design, construction and operation, except the following:

1. Mechanical mixing instead of aeration/mixing is used in IX-SBR process.
2. Ion exchange resins instead of microorganisms are used in IX-SBR.
3. Ion exchange resins are regenerated and recycled to IX-SBR for further treatment (instead of recycling part of microorganisms to bioreactor in conventional SBR).

One of many applications of IE-SBR process is for groundwater treatment which is very similar to the magnetic ion exchange process used at the Village of Palm Springs, FL (22). The difference is that the newly developed IX-SBR process (2) is a batch process, whereas the Palm Springs Plant is using a continuous process. The batch IX-SBR process is equally effective.

1.11. GAC-SBR Processes

There are two kinds of GAC-SBR systems: (a) physicochemical GAC-SBR system; and (b) physicochemical and biological GAC-SBR system. In the physicochemical GAC-SBR system (or PC-GAC-SBR system), there are no microorganisms (biomass, or activated sludge) involved in creating any biochemical reactions. In the physicochemical and biological GAC-SBR system (or PC-BIO-GAC-SBR system), microorganisms (biomass, or activated sludge) are present in the reactor, and the GAC is called biological GAC (25, 35, 36). The PC-GAC-SBR can be used for either potable water purification or wastewater treatment, whereas the PC-BIO-GAC-SBR is still in the research stage for wastewater treatment.

Sirianuntapiboon (36) investigated the efficiencies of both the PC-BIO-GAC-SBR system and the PC-BIO-GAC-SBR system for removal of organic and phenol compounds from the wastewater. The wastewater contained glucose (3500 mg/L COD) and phenol (1000 mg/L). Their results show that the absorption ability of GAC in the PC-BIO-GAC-SBR system was higher than PC-GAC-SBR system. The maximal phenol and COD absorption abilities of GAC in the PC-GAC-SBR system were 213 and 685.1 mg/g of GAC, respectively, although

the maximal phenol and COD absorption abilities of GAC in the PC-BIO-GAC-SBR system were 240 and 761 mg/g of GAC, respectively. The suitable amount of GAC which was supplemented in the PC-BIO-GAC-SBR system was 1000 mg/L.

Comparison of the removal efficiencies between PC-BIO-GAC-SBR system and conventional biological SBR system in various HRT values was made. At the condition of 1 day HRT, the COD and phenol contents in the effluent from PC-BIO-GAC-SBR system were 16 and 0.2 mg/L, whereas the COD and phenol contents in the conventional SBR system were 122 and 3.1 mg/L, respectively. In the conventional biological SBR system, at HRT values of 3, 5 and 10 days, the COD contents of the effluent were 48, 32 and 16 mg/L, respectively. And the phenol contents of effluent were 0.35, 0.18 and 0.018 mg/L, respectively. In the PC-BIO-GAC-SBR system, at the HRT of 3, 5 and 10 days, the COD contents of effluents were 16, 14.5 and 8 mg/L, respectively, the phenol contents of effluents were 0.08, 0.023 and 0.018 mg/L, respectively. From all of the results above, GAC had the advantage for removal of organic matters and toxic substances (phenol compounds) in the SBR system. The mechanism of the GAC in the SBR were adsorption of the organic matters and phenol compounds, the matrix for microorganisms to attach for increasing of the mixed liquor suspended solids (MLVSS) in the system and provided good conditions for microorganisms, attached on the GAC to degrade the phenol compounds. An engineering solution for GAC separation and reuse is needed.

1.12. PAC-SBR and PACT-SBR Processes

There are two kinds of PAC-SBR systems: (a) physicochemical PAC-SBR system; and (b) physicochemical and biological PAC-SBR system.

In the physicochemical PAC-SBR system (or PC-PAC-SBR system), no microorganisms (biomass, or activated sludge) are involved in creating any biochemical reactions, and PAC is dosed to chemical flocs for either water or wastewater treatment, it is the PC-PAC-SBR process (25).

In the physicochemical and biological PAC-SBR system (or PACT-SBR system), microorganisms (biomass, or activated sludge) are present in the reactor, and PAC is dosed to activated sludge mixed liquor for wastewater treatment (25, 26, 35). Both processes are well established, and PACT-SBR has been used in full scale (26).

1.13. VSB-SBR and VSD-SBR Processes

Vertical shaft bioreactor (VSB), also commercially known as deep shaft process or Vertreat process, is an activated sludge process involving the use of a long shaft bioreactor (100 to 500 ft. in depth, and 2.5 to 10 ft in diameter) as the aeration tank for biochemical reactions. (37). The VSB process can be operated as a continuous process or a batch process. Similarly vertical shaft digestion (VSD), also commercially known as Vertad process, is an aerobic digestion process for biosolids treatment. VSD involves the use of a long shaft bioreactor (100 ft in depth and 2.5 to 10 ft. in diameter) as the digester, and can be operated as a continuous process or as a batch process.

When VSB and VSD are operated as batch processes, they are VSB-SBR wastewater treatment process and VSD-SBR sludge digestion process, respectively.

1.14. Physicochemical Membrane-SBR Process

Biological membrane-SBR process which involves the use of a sequencing batch bioreactor for equalization, bio-oxidation, nitrification, denitrification and clarification is introduced in Section 1.6. This section introduces a physicochemical membrane-SBR process (PC-membrane-SBR) in which a sequencing batch chemical reactor is used for chemical feeding/mixing/flocculation/clarification, and the clarified effluent goes through a membrane module for final polishing. When the microfiltration (MF) membrane and the ultrafiltration (UF) membrane are used, the process names will be PC-MF-SBR and PC-UF-SBR, respectively.

1.15. Biosolids SBR Digestion Process

Section 1.9 introduces a solid waste SBR digestion process. This section introduces a SBR digestion process designed specifically for biosolids digestion. At present, many small municipal wastewater treatment plants use batch aerobic digestion process, which in principle, is a biosolids SBR digestion process, with at least the following steps:

1. FILL
2. REACT (AEROBIC DIGESTION)
3. SETTLE (GRAVITY THICKENING)
4. DRAW
5. IDLE

Only the well-established traditional aerobic-anoxic sequencing batch reactor processes (hereinafter referred to as SBR) is presented in the remaining sections of this book chapter (Sections 2 to 6). The traditional SBR is a batch suspended growth activated sludge process involving the use of sedimentation for clarification. A few selected nontraditional SBR processes are presented in Section 7, DESIGN EXAMPLES.

2. TRADITIONAL SBR PROCESS SYSTEMS

2.1. Traditional SBR Process Description

Traditional sequencing batch reactors (SBR) actually represent a very elementary form of treatment process known as fill and draw. Wastewater is added to a reactor, is treated to remove the undesirable components, and is subsequently discharged. The SBR reactor is a self-contained treatment system incorporating equalization, aeration, and clarification within the confines of a single basin. SBR is best operated in an ordered sequence. Single basin operation can provide totally acceptable results, depending on treatment objectives.

The SBR system uses the latest biomass conditioning technology. This enables the SBR system to attain nutrient control without the addition of chemicals and to outperform continuous flow through systems. These advantages, along with technology advancements in hardware devices, are the primary reasons for the increased interest in the use of SBR systems.

A traditional SBR process is similar to an activated sludge process. The main advantages of a sequencing batch reactor process are:

1. Improved effluent qualities,
2. The elimination of separate clarifiers and sludge return pumps,

3. Increased settling area,
4. A perfectly quiescent settling environment,
5. Demand controlled energy consumption,
6. Short-circuiting eliminated,
7. A special ability to handle extremely high organic and hydraulic shock loads, and
8. The capability to equalize flows and load.

Improvements in the application of the SBR process have created a renewed awareness of its advantages (38–44).

The biological SBR can treat a wide range of domestic and industrial wastewater, at flows ranging from a few thousand gallons to millions of gallons per day. (1 gallon = 3.785 liters).

The SBR is unique in its ability to act as an equalization basin, aeration basin and clarifier within a single reactor. The termination of flow and aeration during the treatment process provides perfectly quiescent settling conditions in the reactor and permits even very fine particles to settle. Each reactor maintains its own treatment regime and all phases of treatment occur in each reactor.

As introduced previously, optimum performance is attained when two or more reactors are used in a predetermined sequence of operation. In certain instances, a single reactor may be operated to provide an acceptable level of treatment.

The ratio of raw wastewater influent flow to biomass is a key factor in obtaining desired effluent quality results in the SBR. Because only a small amount of sludge is wasted each cycle, the high quality of the biomass is always maintained.

A true batch biological reactor system, such as the SBR, does not allow influent wastewater to enter the SBR reactor during the final aeration, settle and decant phases, thereby assuring an excellent quality of final effluent.

Simultaneous nitrification and denitrification can be accomplished in SBR process system (17). Combined physicochemical treatment (such as physicochemical SBR) and biological treatment (such as biological SBR) may significantly improve the overall waste treatment efficiency (44).

2.2. Traditional SBR Compared to Other Biological Treatment Systems

Sequencing Batch Reactor systems represent a variation of the activated sludge process. Like any other activated sludge process, the SBR works by developing a mixed culture of bacteria which is effective in removing BOD, COD and nutrients commonly found in wastewater. Other biological treatment systems are of continuous flow operation. The SBR is operated in a true batch reactor treatment mode which does not allow wastewater to enter the reactor during the REACT, SETTLE, and DECANT phases.

The advantages of the sequencing batch reactor over conventional flow-through systems can be compared to the advantages of a typical activated sludge process over a trickling filter process.

With a trickling filter process, the ability to precisely control the degree of treatment is almost impossible. The only controllable option is the wetting rate, i.e. the flow rate over the media. Atmospheric and environmental conditions that exist within the reactor media limit the ability to attain consistent results. Whereas, in an activated sludge process, the food to mass

(F:M) ratio, the dissolved oxygen concentration and MLSS inventory can be monitored and controlled.

The sequencing batch reactor process improve upon this technology. Although flow and load changes can often adversely impact the operation of continuous flow-through systems, the batch system isolates each phase, preventing any adverse effect of these changes. In addition, the environmental conditions can be adjusted to control nutrients within the same reactor.

In comparison with other biological treatment process systems, traditional SBR may have the following advantages:

1. Tolerates Substantial Organic Shock Loads
   Because the SBR reactor serves as an equalization basin during the FILL phase, it can easily tolerate high peak hourly flows and/or substantial organic shock loads without degradation in effluent quality. In fact, small continuous flow activated sludge systems subjected to excessive diurnal variations can show significant improvements in performance when converted to the SBR process.
2. Availability of Phase Management
   Because the treatment of waste is by phases, it is possible to manage each phase so it meets specified requirements. Any of the phases can be increased or otherwise modified to attain desired effluent quality.
3. Resists Solids Washout
   Mixed liquor solids cannot be washed out by hydraulic surges because they can be held in the tank as long as necessary.
4. Pumping Requirements Reduced
   No pumping is required because the mixed liquor is always in the reactor.
5. Ideal Quiescent Setting Conditions
   Solid-liquid separation occurs under ideal quiescent conditions. During the SETTLE phase, short-circuiting is non-existent. Because the settle area is the same as the reactor area, low surface settling rates are achieved, resulting in settling of even small floc particles.

   Aeration Efficiency Intensified
   Because the dissolved oxygen level is zero during the initial FILL phases, a greater oxygen driving gradient exists during the REACT phase, thus achieving higher overall oxygen transfer efficiency with the same aeration applied.
6. Filamentous Growth Eliminated
   Filamentous growth can be controlled by varying the operating strategies during the FILL phase. Filamentous organisms need an oxic condition to survive but by providing an anaerobic or anoxic condition during the FILL phase, the filaments are eliminated. In some cases, extremely high SVI values in a flow through activated sludge system can be reduced to less than 80 using a batch reactor system.
7. Nutrient Removal Without Chemical Addition
   A SBR system can be operated to achieve nitrification, denitrification, or phosphorus removal without chemical addition. Nitrification can be achieved by increasing the duration of the REACT phase or the mixed/aerated portion of the FILL REACT phase. Denitrification can be achieved by increasing the length of the STATIC FILL or MIXED FILL phases so that zero or near zero dissolved oxygen conditions exist during these period and the incorporation of anoxic segments reducing the FILL REACT and REACT phases.

Phosphorus removal is accomplished by selecting a control strategy which eliminates oxidized nitrogen and dissolved oxygen during the STATIC FILL phase or MIXED FILL phase and allows for aeration during the FILL REACT and REACT phases.

Phosphorus release takes place during the STATIC FILL and MIXED FILL phases when oxidized nitrogen and dissolved oxygen are eliminated. This is followed by luxury phosphorus uptake during the REACT phases when aerobic conditions occur in the reactor. These variations in operating strategies are unique to the SBR systems and can be easily achieved by simple adjustments in the microprocessor control.

9. Small "Foot Print"

Because the SBR is operated in a true batch treatment mode, optimum effluent quality is obtained during each cycle. Only a fraction of the total reactor volume, typically 1/6, is introduced into the reactor each cycle. This raw flow combines with the acclimated biomass, which remains in the reactor at all times.

3. PRINCIPLES AND OPERATION OF TRADITIONAL SBR PROCESS

3.1. Process Principles

The traditional SBR process operates on a Fill and Draw principle. Normally, the process follows the basic steps of Fill, React, Settle, and Decant. The actual cycle time will vary with the effluent results desired. If only BOD reduction is desired, a cycle time as short as 3 hours may be used. If further treatment to obtain nutrient control is required, the cycle time can be extended to accommodate the process requirements.

Process chemistry and theory and principle of biological SBR are identical to that of batch activated sludge process (38–44), therefore will not be repeated here. The readers are referred to the chapter of activated sludge process in this handbook.

3.2. Operational Phases

Figure 11.1 depicts the various traditional SBR phases.

The FILL phase can include many phases of operation and is subject to various modes of control. Perhaps the earliest (and most elementary) mode of control of the FILL phase is based on reactor liquid level or volume. Using this method, the FILL phase is terminated when some preselected volume or depth has been attained. It is apparent that this approach provides cycle times that are inversely related to flow rates. If peak flow rates are the basis for design, then longer cycles will occur under lower flow conditions, resulting in over aeration, excessive use of energy and potential degradation of biomass. Other choices of control can include the use of plant flow rate or the preferred and proven choice of "time" to dictate the FILL phase duration.

Depending on the treatment requirement of the specific SBR design, the FILL phase may be composed of STATIC FILL, MIXED FILL, as well as REACT FILL increments.

Under STATIC FILL, influent flow is introduced to the reactor under a nonmixed, nonaerated environment. This phase may be necessary or helpful under operating strategies devoted to nutrient control or in minimizing energy requirements.

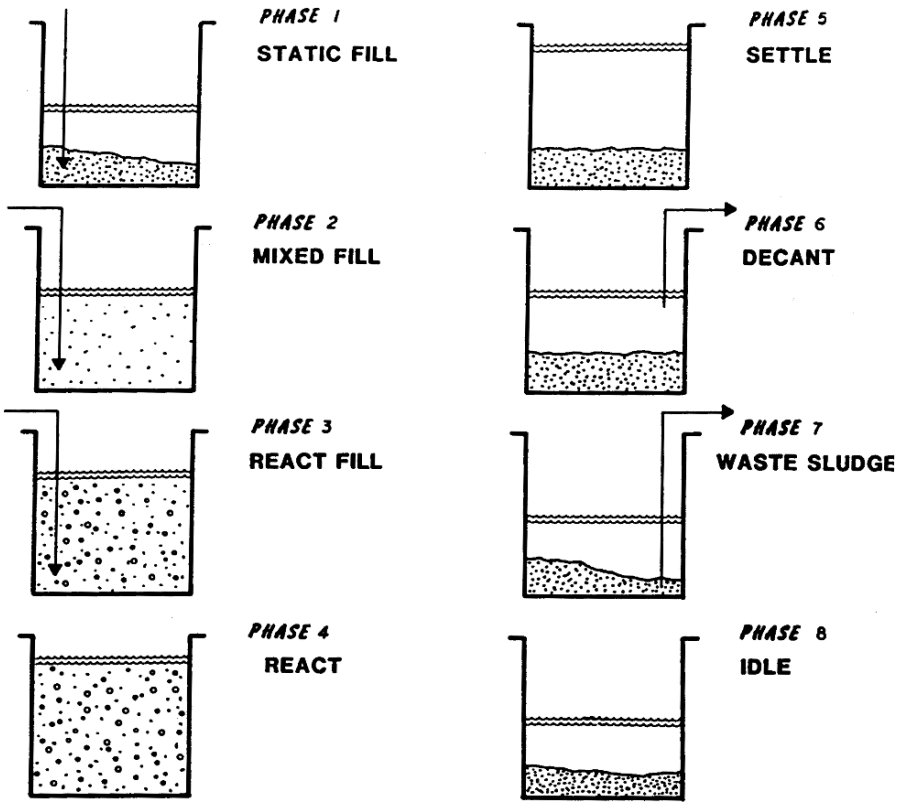

Fig. 11.1. Various operating phases of traditional biological SBR system.

Closely related to this phase of operation is the MIXED FILL phase. As the name implies, this phase provides reactor mixing without aeration, which creates either an anoxic or anaerobic environment.

REACT FILL is that increment of the FILL phase accompanied by both mixing and aeration.

REACT identifies the completion of the REACT phase that occurs after FILL has been completed. With a single reactor, continuous inflow system, there is no separation between the REACT and the REACT FILL phases.

The REACT phase is typically time measured, but other methods of control, such as the measurement of specific substrate removal, are also possible. Sometimes incorporated within the REACT phase is a very short mix only (no aeration) period of 2 to 5 minutes. It has been found that, during initial start up periods of winter (cold weather) operation, a mix only period just before settle enhances the flocculation and improves settling.

SETTLE represents the quiescent phase during which no aeration or mixing occurs, normally it is time controlled.

DECANT is the phase during which clarified effluent is withdrawn from the basin and it is usually controlled by level.

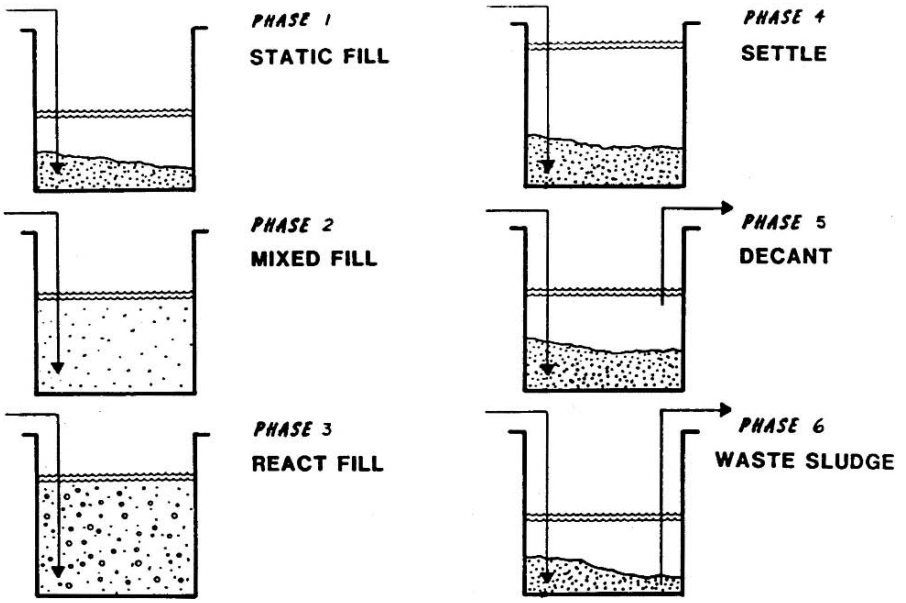

Fig. 11.2. Operating phases of a single basin continuous inflow SBR system.

The final phase, IDLE, is used only when there is more than one basin. This phase allows the reactor to remain idle until the FILL phase is completed in the basin being filled. There is no IDLE phase with a single basin continuous inflow system. A single basin sequence is depicted in Figure 11.2.

3.3. Food to Microorganism Ratio (F:M)

Although the F:M ratio in a traditional SBR system continually changes owing to variances during the FILL and REACT phases, the average values are comparable to those in activated sludge systems. Currently, most proposed SBRs are planned around a low load design, typically with a F:M ratio of 0.05 to 0.1. This is comparable to an extended aeration type process.

Conventionally loaded activated sludge plants are operated in the range of 0.15–0.4 F:M. Sequencing batch rectors can also be designed to operate in this range. When nutrient control is required, a lower F:M is required (39, 41).

High rate systems, with F:M rates of 1.0 or higher, are also possible and are particularly useful for industrial pretreatment applications.

W.M. Shubert (1) and K.A. Mikkelson (4) recommend the following SBR design parameters:

1 For municipal wastewater treatment low load design:
   a. Food to microorganisms ratio 0.05–0.1/d
   b. Treatment cycle duration 4.8–6 hours
   c. Typically low water level MLSS 4000–4500 mg/L
   d. Hydraulic retention time 18–24 hours

2. For municipal wastewater treatment conventional load design:
   a. Food to microorganism 0.15–0.4/d
   b. Treatment cycle duration 4 hours
   c. Typically low water level MLSS 2000–2500 mg/L
   d. Hydraulic retention time 6–14 hours
3. For industrial wastewater treatment low load design:
   a. Food to microorganism 0.05–0.1/d
   b. Treatment cycle duration 4–48 hours
   c. Typically low water level MLSS 4000–6000 mg/L
   d. Hydraulic retention time variable with waste concentration
4. For industrial wastewater treatment conventional load design:
   a. Food to microorganism 0.15–0.6/d
   b. Treatment cycle duration 4–24 hours
   c. Typically low water level MLSS 2000–4000 mg/L
   d. Hydraulic retention time variable with waste concentration

4. PROCESS APPLICATIONS

Traditional SBR technology permits organic removal, including nutrient removal or conversion, all within the confines of a single reactor.

4.1. BOD Reduction

The removal of organics, usually measured as BOD_5, occurs throughout the SBR process. Most of the bacteria in the reactor are facultative in nature. They will remove BOD_5 anaerobically when oxygen is absent (FILL phase) and aerobically when oxygen is present (REACT phase). Oxygen requirements are calculated assuming only aerobic removal of BOD_5. Because only part of the cycle is devoted to aeration, oxygen demand must be satisfied within this interval. The required cycle time for BOD_5 removal only is less than that for operation to control nutrients.

4.2. Nitrogen Removal

The oxidation of ammonia to nitrates (nitrification) involves two steps: conversion of ammonia first to nitrite and then to nitrate. The bacteria responsible for this conversion are aerobic and the oxygen requirement is approximately 4.6 lbs of oxygen per lb. of ammonia nitrogen applied.

When the maximum nitrogen removal is required, nitrates (and nitrites) can be reduced through biological actions.

In the absence of dissolved oxygen (DO), several facultative bacteria commonly present in wastewater treatment systems are able to use nitrates as a terminal electron acceptor with the resultant formation of nitrogen gas (denitrification). The electron donor (carbonaceous energy source) is usually present as a natural carbon source in the wastewater in an SBR basin.

The necessary conditions must be present to achieve nitrification and/or denitrification. The conditions for nitrification are sufficient DO and alkalinity, plus a carbon source. Denitrification requires near zero DO, a carbon source, and nitrates. In addition, the conditions in the reactor must be appropriate for maintaining the nitrifying and denitrifying bacteria.

To ensure the growth of nitrifying organisms, it is necessary to provide a sufficient solids residence time and a suitable aerated basin volume at a dissolved oxygen level adequate for nitrification to occur. An anoxic period in the SBR cycle is required for nitrate reduction. Excessive aeration times and high DO levels promote nitrification at the expense of the carbon source for nitrate available for denitrification. Alternately, the anoxic conditions necessary for denitrification provide an environmental in which the nitrifying organisms cannot grow. Also, as conditions are changed to allow for more and more unmetabolized carbon source to be present to accelerate the denitrification rate, there will be a corresponding increase in the concentration of ammonia which must first be oxidized to nitrates to its availability for reduction. These facts require that the cycle increments be carefully controlled.

4.3. Phosphorus Removal

Phosphorus removal can be accomplished by conventional methods such as the addition of a suitable coagulant, such as ferric chloride, and precipitating the phosphorus out via the sludge. However, this method of phosphorus removal will produce substantially greater volumes of sludge and will have an adverse effect on the overall operation of the system.

With the SBR process, biological removal of phosphorus can be achieved without chemical addition.

The biological removal of phosphorus first requires an anaerobic period (i.e., the absence of dissolved oxygen and oxidized nitrogen) and the presence of exogenous electron donors (i.e., the substrate) during the anaerobic period. This period should be followed by an aerobic period (presence of dissolved oxygen), which promotes luxury uptake of phosphorus by the biomass. A suitable quantity of the sludge mass is removed from the reactor in the waste sludge phase before the next anaerobic period. The control strategy eliminates oxidized nitrogen and dissolved oxygen during part of the FILL phase (anaerobic period) and provides aeration during REACT.

It is possible to attain biological phosphorus reduction of 90%. However, a realistic design is to plan biological reduction down to 2 to 3 mg/L. If the requirement is less than the 2 to 3 mg/L, chemicals, such as alum, can be added to the reactor or to the effluent, followed by filtration to obtain further reduction. The amount of chemical needed for this addition is small. The solids generated are backwashed to the sludge holding tank or digester. Because the SBR process provides for biological phosphorus removal, chemical and sludge handling costs are greatly reduced.

4.4. Municipal Domestic Applications

SBR systems are currently used in a wide variety of treatment situations, under a wide range of loading conditions. The SBR process can be employed to treat any wastewater that can be treated by activated sludges system, including: anoxic/oxic systems, aerated lagoons or extended aeration systems, trickling filters, RBC, membrane bioreactor systems and oxidation ditches.

Because SBR has high efficiency for removal of BOD/COD, TSS, nitrogen and phosphorus, it has been widely used for treatment of wastewater from the municipalities, casino, resorts and institutions (see Practical Examples).

4.5. Industrial Applications

SBR has been successfully applied to the treatment of the following industrial wastewater:

1. Chemical and petrochemical
2. Leachate
3. Tobacco
4. Food
5. Pulp and paper
6. Dairy
7. Beverage
8. Textile
9. Tannery
10. OCPSF (organic chemical, plastics, synthetics and fibers)
11. High nitrogen
12. Any other wastewater that can be biologically treated.

5. PROCESS DESIGN

5.1. Flow and Cycle Time

The normal operating cycle time of a traditional SBR low load design to attain BOD removal and nutrient control is 4.8 hours. Each cycle creates several different environments and consists of eight specific phases in a multiple basin configuration (11).

The overall fill period is normally time rather than level controlled. Sufficient volume is allocated to handle maximum daily flow. Flow rates at less than this rate result in termination of the FILL phase at less than maximum capacity of the reactor. At flows greater than the maximum daily rate, a level sensing device terminates the FILL phase.

During the early life of a facility, when the flow is significantly less than design capacity, the time control maintains the preset cycle, using a fraction of the SBR basin capacity. The fixed time interval ensures complete organic removals. Aeration controls, such timers or DO monitors to control the operation of the aeration devices, prevent wasted power and over-aeration.

Some engineers and manufacturers use fill time as a function of the plant flow rate as a basis for design. This does not provide for the repetitive controlled environment to permit consistently high quality effluent. Time control should be the basis of design; and this time increment should be constant at all flows up to maximum daily design flow.

5.2. Process Phase Design

The first phase is the STATIC FILL phase and represents a specific field adjustable time period. At flow rates higher than design, the sensing of an intermediate level overrides the time control.

The second phase is a MIXED FILL phase, which also represents a specific time increment. During his period, the reactor contents are thoroughly mixed with a mixer that is capable of suspending all biological solids without aeration. Again, level sensing is used to override time at greater than maximum design flow rates.

The third phase is the REACT FILL phase and represents a specific rise in level in the reactor basin (established at the maximum daily flow rate) and/or a specific field adjustable time period. This phase is usually DO or time controlled to maintain the dissolved oxygen level at the predetermined field adjustable level.

The sum of the various FILL, increments described in the preceding paragraphs occupy 144 of the 288 minutes of the complete cycle when operated at flows of equal or less than maximum daily flow. Flows in excess of maximum daily flow are level controlled, which shortens the cycle.

The fourth phase is the REACT phase. During this phase, there is no influent flow, and this is the time when the final portion of the react period takes place. The level at his point remains constant. The time period is field adjustable, but normally allocated at 35 minutes. Also refer to Section 3.2.

The fifth phase is a SETTLE phase. The level remains constant and is set for a field adjustable time period, usually 45 minutes. This quiescent environment promotes settling that surpasses the performance of a conventional clarifier.

The sixth phase is the DECANT phase. This is the time when the treated effluent is removed, and the level is reduced from the maximum fill level to a predetermined low level. The time element associated with this varies, depending on how far the basin is filled within the time allocated for the fill period. The decant rate is established to reduce reactor volume from maximum to minimum level within 60 minutes. Because the decant rate is constant, decant time will vary with volume.

The seventh phase is the WASTE SLUDGE phase and lasts for a field adjustable time period. The level will remain approximately at the low water level. The WASTE SLUDGE phase can be incorporated into the last portion of the DECANT phase.

Sludge can be wasted at any interval the operator selects. The operator can waste settled sludge at the end of or during the DECANT phase, or he can waste mixed liquor at or near the end of the REACT phase. However, the most logical time to waste sludge, and by far the easiest to manage, is after or near the end of the DECANT phase.

The eighth phase is the IDLE phase and can last anywhere from 0 to 60 minutes, depending on the cycle time of the other basin or basins.

Figure 11.3 illustrates graphically the traditional SBR cycles under average daily flow conditions for a dual reactor system.

Note that FILL is continuous. When Basin #1 reaches the allotted fill time, flow is switched to Basin #2. The REACT, SETTLE and DECANT phases may then progress in Basin #1. Because the decant rate is adjusted to remove the maximum daily flow in the allotted decant time period, idle time results until Basin #2 reaches its allotted time period. The idle time approaches zero as the flow rate for a particular cycle approaches the maximum daily rate. This condition is illustrated in Figure 11.4.

At peak flow rates greater that the maximum daily rate, the process controller of the SBR system receives liquid level inputs from level sensing devices and automatically adjusts the cycle time to permit maximum settling and decant efficiencies.

The typical nominal operating cycle for the traditional SBR design is detailed in Table 11.1.

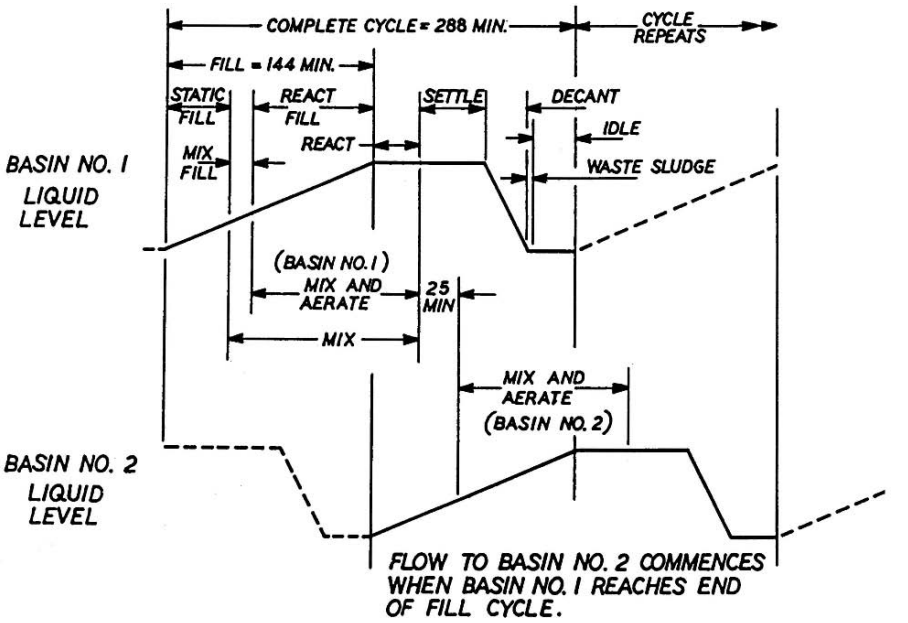

Fig. 11.3. Two basin SBR process cycle at SBR design flow (1).

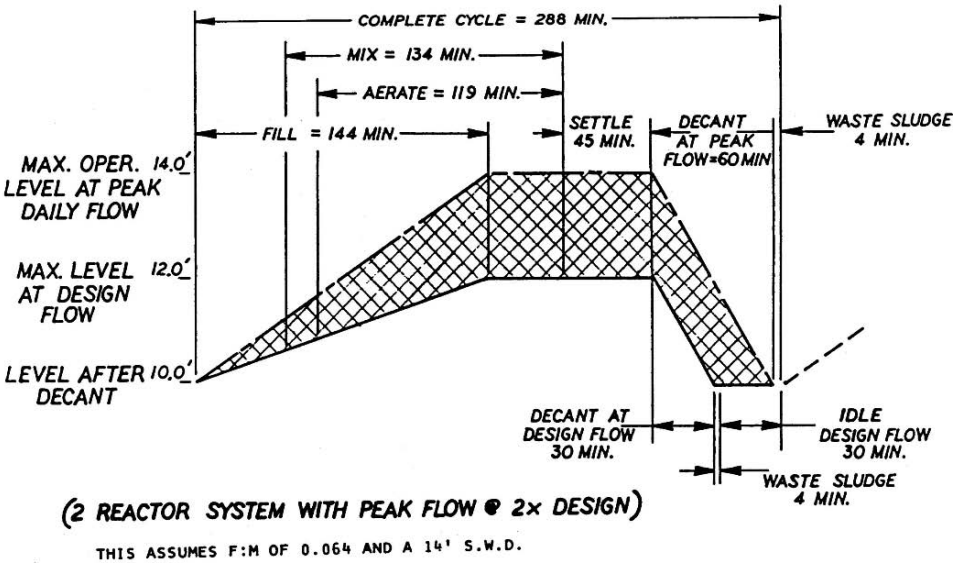

Fig. 11.4. Effect of influent flow rate on SBR cycle time and basin operating levels (1).

Table 11.1
SBR multiple basin operation for BOD and nutrients removals

Phase	Environment	Water level (elevation), ft		Influent value	Mixer	Aeration blowers	Decant system	Waste sludge pump	Typical times/field adjustable	Notes
		Start	End							
1 Static Fill	Anoxic	10.0	11.25	Open	Stopped	Stopped	Stopped	Stopped	45 minutes	
2 Mixed Fill	Anoxic	11.25	11.67	Open	Operating	Stopped	Stopped	Stopped	15 minutes	
3 React Fill	Oxic/Anoxic	11.67	14.0	Open	Operating	Operating	Stopped	Stopped	84 minutes	Blowers are programmed On or Off to maintain DO
4 React	Oxic	14	14	Closed	Operating	Operating	Stopped	Stopped	35 minutes	
5 Settle	—	14	14	Closed	Stopped	Stopped	Stopped	Stopped	45 minutes	
6 Decant	—	14	10.0	Closed	Stopped	Stopped	Operating	Stopped	30 minutes Avg. Flow* 60 minutes Max. Flow*	
7 Sludge Waste	—	10.0	10.0	Closed	Stopped	Stopped	Stopped	Operating	4 minutes	
*8 Idle	—	10.0	10.0	Closed	Stopped	Stopped	Stopped	Stopped	30 minutes Avg. Flow 0 minutes Max. Flow	

179 minutes fill & react

STATIC FILL – (No Mix) or (No Air)
MIXED FILL – (Mix, but no Air)
AERATED FILL – (Mix and Air)
REACT – (Mix and Air)
SETTLE – (No Mix) or (No Air)
DRAW – (No Mix) or (No Air)
WASTE SLUDGE – (No Mix) or (No Air)
IDLE – (No Mix) or (No Air) – Applies only to 2 or more reactors

*Maximum to Average flow assumed to be 2:1. Interval is level, rather than time controlled.
Aeration time is normally 1.58 hours per cycle.
Mix time is normally 2.23 hours per cycle.

5.3. Process Modifications

To attain various degrees of treatment, specific phases are added or deleted. Although the SETTLE and DECANT phases may remain constant, the FILL and REACT phases are adjusted to meet the various conditions needed to attain the desired biological environment.

1. BOD reduction only (Refer to Table 11.2: TSS 30 mg/L and BOD 30 mg/L effluent
   Where nutrient control or removal is not required, the cycle time is usually shortened to three hours. Table 11.2 reflects the cycle increments for this operating mode.
2. BOD and phosphorus reduction (Refer to Table 11.1)
   Use Phases #1 or #2, #3, #4, #5, #6, #7, and #8.
3. Nitrification, denitrification for near total nitrogen reduction, as well as BOD and phosphorus reduction (Refer to Table 11.1).
   Use Phases #1, #2, #3, #4, #5, #6, #7, and #8.

To increase react time, overall length of the operating cycle does not necessarily have to be increased correspondingly. Longer react time required for nitrification, or for the treatment of high strength waste, can be achieved by running the aeration equipment for a larger portion of the FILL phase. This mode of operation may also be used when the design capacity has been exceeded in an existing plant and the additional flow has to be treated in the existing SBR basin, temporarily or permanently, as the need may be.

The standard design includes aeration for a portion of the FILL phase to promote the maximum treatment level at all times. Aeration can be controlled by time or dissolved oxygen. If nitrification is not desired, field adjusted cycle time can be shortened, or the STATIC FILL phase and MIXED FILL phase can be lengthened to reduce the react time. MLSS concentration (and the resulting F:M ratio) can be adjusted by varying the WASTE SLUDGE phases to attain the desired degree of treatment.

Operating strategies for nitrification and denitrification may not necessarily be different. Recognizing that nitrification must precede denitrification, identical operating strategies can be expected if the DO is reduced to near zero during SETTLE, DECANT and IDLE phases.

Because phosphorus removal requires an anaerobic period followed by an aerobic period, a denitrifying system is easily adaptable to the removal of phosphorus, provided sufficient incoming BOD is available as a carbon source for the denitrifying organisms.

The basin will stratify during the SETTLE and DECANT phases and will remain stratified during the STATIC FILL phase, which will permit zero DO (anaerobic) conditions to exist at the bottom of the basin while maintain aerobic conditions in the upper level. This allows for an anaerobic condition to exist without causing an odor problem.

5.4. Decanter System Design

The decant system is the heart of the SBR.

It must be designed to prohibit mixed liquor suspended solids from entering the decanter during the MIXED or REACT phases. Any MLSS that enter the decanter during these phases will settle in the bottom of the decanter and will be flushed out during the first few minutes of the DECANT phase. This high concentration of solids that can exceed the MLSS concentration will have a serious adverse effect on effluent quality. Therefore, an absolute seal

Table 11.2
SBR multiple basin operation for BOD removal only

Phase	Environment	Water level (elevation), ft Start	End	Influent value	Mixer	Aeration blowers	Decant system	Waste sludge pump	Typical times/field adjustable	Notes
1 Mixed Fill	Anaerobic	10.0	10.7	Open	Operating	Stopped	Stopped	Stopped	16 minutes	
2 React Fill	Oxic/Anoxic	10.7	14.0	Open	Operating	Operating	Stopped	Stopped	74 minutes	Blowers are programmed On or Off to maintain D.O.
3 React	Oxic	14	14	Closed	Operating	Operating	Stopped	Stopped	16 minutes	
4 Settle	—	14	14	Closed	Stopped	Stopped	Stopped	Stopped	30 minutes	
5 Decant	—	14	10.0	Closed	Stopped	Stopped	Operating	Stopped	22 minutes Avg. Flow 44 minutes Max. Flow	
6 Sludge Waste	—	10.0	10.0	Closed	Stopped	Stopped	Stopped	Operating	(4 minutes)*	
7 Idle	—	10.0	10.0	Closed	Stopped	Stopped	Stopped	Stopped	22 minutes Avg. Flow 0 minutes Max. Flow	

Maximum to average flow assumed to be 2:1.
Aeration time is normally 1.50 hours per cycle.
Mix time is normally 1.77 hours per cycle.
TOTAL CYCLE TIME = 3 hours

* *Note*: Waste Sludge commences 30 minutes after initiation of decant. Time is included in either Decant or Idle phase.

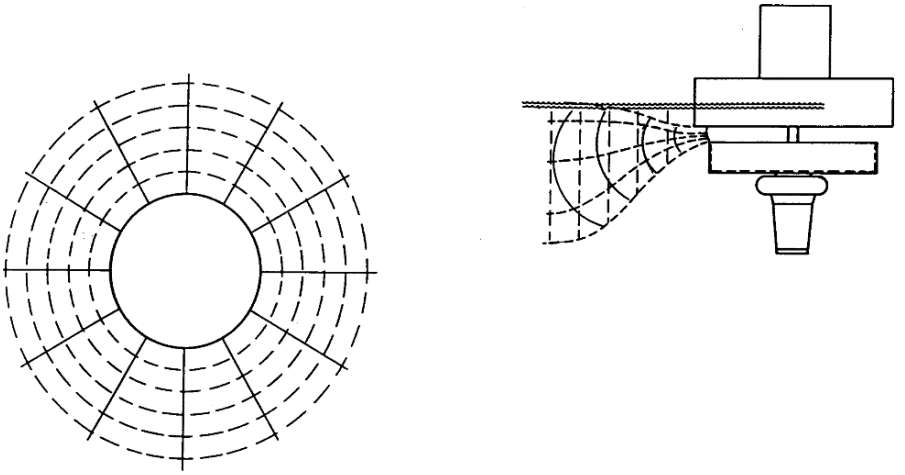

Fig. 11.5. Circular Weir of SBR decant system (1).

at the decanter entrance interface is a necessity. Some designs on the market provide air seals or flapper valves. These do not provide the necessary absolute seal needed to assure repeated effluents free of solids. The only way of assuring that an absolute separation is obtained is to provide a reliable seal as offered by a reliable and reputable SBR manufacturers, or totally remove the decant system from the reactor except during the DECANT phase.

Decant systems that do not provide an absolute seal, or those that are not removed from the reactor during the REACT and MIXED phases, must be designed with a flush system that recirculates the first several minutes of decant liquid.

The decanter should be designed to follow the liquid level down during the DECANT phase permitting short settle periods. It should also be designed to draw clear effluent from 6 to 8 in. below the surface, prohibiting floating scum from entering the decant system. Proper decanter design ensures clearer effluent even when sludge volume indexes (SVIs) are in the 200+ range. Fixed decanter designs should be avoided because operation options are reduced and obtainable effluent quality is limited.

Because entrance turbulence can pull lighter sludge up and seriously affect effluent quality, entrance velocities must be considered in the decanter design. The circular weir and decant system provide the optimum in energy dissipation. Although the velocity at the weir entrance may be high, the velocity dissipates very quickly owing to the increased affected area as it radiates from the weir (Figure 11.5).

Most other decanters on the market at this time produce flow patterns that limit their capacity owing to the localized high velocities generated from single takeoff points.

Decanters using an air lock system do not provide the necessary mechanical seal and create operational problems. The air is released through an electrically operated solenoid valve. The air has a very high humidity level and will freeze in the winter months unless properly heated and insulated.

The properly designed decanter provides an absolute seal, vertical and horizontal separation of the submerged weir from the basin surface and sludge blanket, as well as a floating support to follow the changing surface elevation during the DECANT phase.

5.5. Skimming System Design

Many biological SBR systems do not incorporate the use of a skimming device because of the varying liquid levels and the complexities that this adds to their design. Operating problems have resulted in some existing systems using competitive aeration approaches, but these are minimized in the traditional SBR design.

The complete separation of the mixing and aeration devices in the traditional approach is responsible for this improvement. The surface down pumping mixer entrains the floating scum and puts it into suspension during the mixing and aeration operating modes. As most of the scum constituents are organic and biodegradable, their accumulation is limited by bacterial action.

Although a thin scum layer does form on the surface during the SETTLE and DECANT phases, this condition does not affect the effluent quality. The design and operation of the decanter prevents this scum from being discharged during the DECANT phase.

Only the traditional SBR system provides this unique combination of equipment that controls scum accumulation.

5.6. Energy Input Optimization

To optimize and/or reduce energy input, there must be a separation of mixing energy and aeration energy. This is obvious from the various phases of the SBR process. Most current biological SBR systems use jet aerators during the MIXED FILL or MIXED phases where no oxygen is introduced into the wastewater. During these phases, the jets are not gassed and only the jet pumps operate. The jet aerators are designed to have the jet motive pumps operate with air input, and elimination of one or the other of these cuts the energy input substantially and serious affects the mixing ability or oxygen dispersion. A jet system designed for oxygen input does not necessarily produce good mixing without air input.

To solve this problem, many jet manufacturers suggest a ditch configuration or a circular mixing motion to take advantage of the propulsion effect. This creates another problem. Liquid in motion in a circular or ditch concept is difficult to stop. The net result is prolonged movement, which in turn prohibits settling and increases the time before the DECANT phase can be initiated.

Absolute separation of mixing and oxygen input can be thereby permitting optimization of both independently. The unique down pumping mixer can provide total biological suspension of solids and oxygen dispersion with less than 25 HP per million gallons of volume. In addition, owing to the unique mixing patterns, the motion terminates vary quickly when the mixer is stopped, thereby promoting fast settling and early initiation of the DECANT phase.

An optimized design permits options in selecting either coarse bubble or fine bubble air diffusion. The use of the down pumping mixer enhances oxygen transfer efficiencies in an aerated basin by as much as 25%, and potentially increases the transfer efficacy in an SBR system by more than 50%. The alpha factor normally associated with a fine bubble diffused

air system is 0.4 to 0.6. Because the down pumping mixer provides the mixing and agitation, the traditional SBR attains an alpha factor of 0.8 or higher. The use of a retrievable fine bubble diffused air system permits attainment of high transfer efficiencies and easy accessibility for cleaning.

The sliming problems associated with fine bubble diffusers in plug flow reactors do not exist in the traditional biological SBR design. The high organic loads existing in the front end of a plug flow design do not occur in the completely mixed reactor. In addition, the diffusers in an SBR system are always gassed at design air flow rate during the aeration phase. Further, a down pumping mixer also provides a scouring action on the surface of the diffusers.

5.7. Three Design Steps

In spite of the phenomenal interest in the biological SBR process, there has not been a unified approach to biological SBR design. Because we are dealing with a form of activated sludge, it follows that some of the same basic design theories can be used. Therefore, MLSS concentration, SRT, F:M ratio, flow rate, and effluent quality all need to be taken into consideration.

The first step is to decide if primary treatment is needed. The only primary treatment suggested for the biological SBR process is a screening device with a 1/4–3/8 in opening to remove the large waste solids present in wastewater. The decision to provide primary sedimentation should be based on the economic feasibility of providing anaerobic digestion to stabilize all organic solid generated. A degrit system may be required if a substantial amount of grit is anticipated in the flow.

The second step to designing a biological SBR process is to establish the effluent quality that must be attained. This, in turn, will determine what F:M ratios will be used, which will establish reactor sizing.

The third step is to determine flow rates–design flow, maximum daily flow, peak hourly flow. The biological SBR process can handle peak hourly flows of 3 to 10 times design flows without affecting effluent quality; but anticipated sustained maximum daily flows must be considered in the design to prevent effluent degradation.

6. SUMMARY AND CONCLUSIONS

6.1. General Summary

The SBR system represents a unique combination of mechanical components and process options that provide design engineers with a highly flexible and versatile basis for designing wastewater treatment systems.

From a mechanical component perspective, the extremely efficient floating circular decanter system represents a major advancements in the overall acceptance of the sequencing batch reactor approach to wastewater treatment.

The key elements of this device include a positive seal in the closure system and the fact that the unit is designed as a floating structure. The positive seal prohibits the entry of suspended solids into the outlet structure during the mixing and aeration modes of operation. This feature permits the discharge of high quality effluent at the initiation of each decant cycle.

Construction of the unit as a floating structure allows for the maintenance of an ideal effluent withdrawal position just below the water surface and well above the sludge blanket throughout the entire decant cycle.

The provision for independent mechanical systems to accomplish mixing and aeration is imperative for designing and operating the biological SBR system. The down pumping mixer capability of efficient independent mixing provides an assurance that:

1. The process reactions will occur in a completely mixed reactor environment.
2. Permits the delivery of an oxygen supply in a completely mixed environment at a rate that closely parallels the system oxygen demand which results in a conservation of energy while optimizing treatment efficiency.
3. Represents the main mechanism for modifying the reactor environment with respect to aerobic or anoxic conditions in biological nutrient removal applications.
4. Allows for a variety of oxygen delivery systems, including surface mechanical aeration and retrievable diffused aeration systems located at the periphery of small or large reactors.
5. Provides systems that are easily maintained without the necessity of dewatering the reactors.

From a process perspective, the SBR represents a variation of the widely accepted activated sludge process. The combining of the typical aeration and solid-liquid separation unit processes within a single vessel with a time basis of operation represents specific advantages to the treatment efficiency of SBR system. The adoption of this mode of operation provides the opportunity to implement specific wastewater treatment strategies within the individual phases of a treatment cycle that are not typically possible in a conventional flow-through activated sludge system.

The incorporation of an ideal solids-liquid separation process is often the key element in achieving a high level of treatment efficiency. The SBR incorporates Settle and Decant phases of operation that are implemented in an ideal quiescent environment that is not subject to the effects of hydraulic factors. This quiescent environmental is maintained at all defined flow conditions to the system.

The combination of highly efficient and reliable SBR mechanical components, coupled with a wide range of process design options presents the design engineer with a very effective basis for the design of wastewater treatment systems. Common applications will range from organic reduction to biological nutrient removal in municipal cases to the pretreatment or complete treatment of a wide range of types of industrial wastewater (4).

6.2. Performance Evaluation

Environmental Canada's Great Lakes 2000 Cleanup Fund, the Water Environment Association of Ontario, and the Ontario Ministry of the Environment sponsored a program to evaluate and document traditional SBR performance in the US Great Lakes region and Ontario, Canada in 1997. Information on the application and performance of traditional SBR technology in 75 municipal plants was obtained from the plant management, and actual site visits. The study results show that traditional biological SBRs can meet quite stringent effluent criteria for organic and nutrient removal. The facilities studied by the investigators (10) met and, in most cases, exceeded their effluent requirements. Table 11.3 presents the surveyed representative results.

Table 11.3
SBRs meet stringent effluent criteria in the Great Lakes Region (10)

Plant/supplier	Actual (design) flow, m^3/d		BOD[1], mg/L	TSS[2], mg/L	Ammonia-nitrogen, mg/L	Nitrate-nitrogen, mg/L	Total phosphorus TP, mg/L	Filtration	Chemical addition
New Freedom, Pa.	4100	Influent	73	81	NA	NA	NA	No	No
	(8520)	Effluent	<5	5	0.8	NA	0.9		
Garden Spot, Pa.	90	Influent	276	380	33	NA	10	Yes	Yes
	(1060)	Effluent	<5	<5	0.4	NA	1.0		
Flushing, Mass.	6880	Influent	120	122	13	NA	2.8	No	No
	(7570)	Effluent	<5	<5	0.5	0.2	0.5		
Soaring Eagle, Mich.	760	Influent	285	190	65	NA	7.0	Yes	Yes
	(2200)	Effluent	<5	<5	0.5	NA	0.2		
Catawba Island, Ohio	1730	Influent	236	394	21	NA	7.5	No	Yes
	(5070)	Effluent	10	12	3.6	0.9	0.5		
Casinorama, Ontario, Canada	700	Influent	289	375	18.3	NA	9.5	Yes	No
	(2100)	Effluent	<4	<5	0.6	0.7	0.3		
Frackville, Pa.	3030	Influent	207	188	NA	NA	5.95	Yes	Yes
	(5300)	Effluent	<5	<5	1.0	5.0	0.5		

1. BOD = *biochemical oxygen demand.*
2. *TSS = total suspended solids.*

Two biological treatment levels have been considered and established:

1. Advanced wastewater treatment with nutrient removal; the effluent limits are:
   (a) $BOD_5 = 10\,mg/L$ (5-days biochemical oxygen demand)
   (b) $TSS = 10\,mg/L$ (total suspended solids)
   (c) $TP = 3\,mg/L$ (total phosphorus)
   (d) $TN = 5\,mg/L$ (total nitrogen)
2. Secondary treatment with nutrient removal, the effluent limits are:
   (a) $BOD_5 = 25\,mg/L$
   (b) $TSS = 25\,mg/L$
   (c) $TP = 3\,mg/L$
   (d) $TN = 5\,mg/L$

Most of the biological SBR plants evaluated met the more stringent of the two limits (10, 11).

6.3. Cost Evaluation

Construction cost data submitted by 17 facilities in the same study sample were compared by the same Canadian investigators (10) with cost estimates provided in the literature. Figure 11.6 shows the unit construction cost (1998 US dollars) as a function of plant capacity for both SBR plant and continuous flow activated sludge plants. This comparison shows that construction cost differences between SBRs and continuous flow activated sludge plant are significant owing to the following main reasons:

1. Lack of need for an extend secondary sedimentation clarifier and return sludge pumping system clearly offers potential saving in SBR construction cost.
2. SBR plants typically do not use primary sedimentation clarifiers. None of the 75 plants evaluated had primary plants.

6.4. Operation Evaluation

With modern automation and instruction, most of SBR plants are in good to excellent operational condition. A few common operating issues for SBR plants have been reported (10):

1. Operators do not have formal training in SBR operation. Professional associations, such as Water Environment Federation produce training manuals for all biological treatment process, except SBR (45).
2. Mechanical equipment – such as air valves, solenoid valves, and decanter arms – located outdoors may freeze or malfunction in the winter.
3. Decanters are not always adequate for specific treatment requirements. For example, decanters may be unable to reduce the discharge of floating material and grease, which clog sand filters and cover ultraviolet-disinfection lamps.
4. Lack of proper aeration control reduces potential energy savings.
5. Inadequate design of pretreatment systems (bar screens, comminutors) may cause accumulation of floating and coarse material in the SBRs, flow metering inaccuracies, and other potential problems.
6. Some SBRs are supplied without a specific sludge-age control strategy.

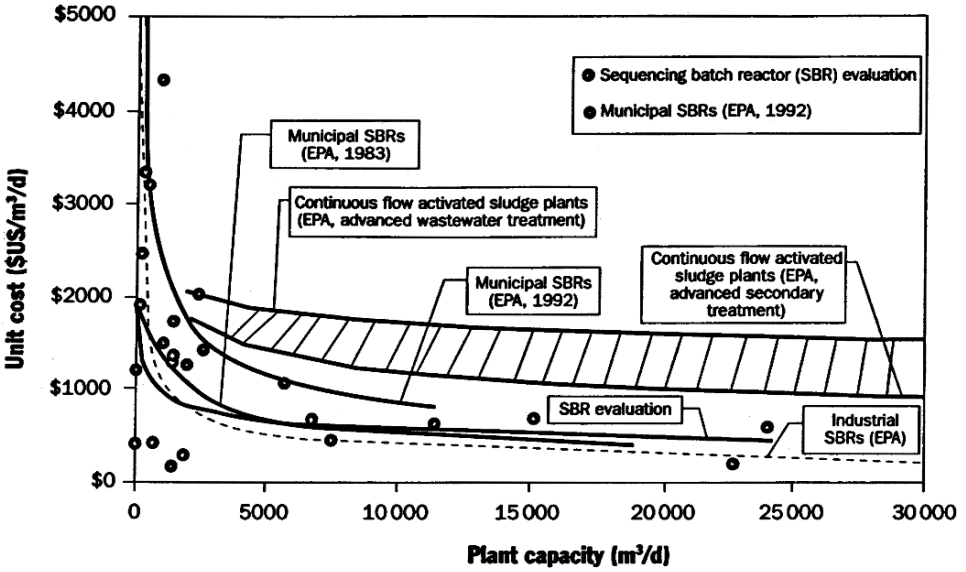

Fig. 11.6. Unit construction cost (1998 US Dollars) as a function of plant capacity (10).

7. Many SBRs use aerobic digesters for sludge treatment. If biological phosphorus removal is performed, a considerable portion of the phosphorus removed in the waste activated sludge may return to the SBRs if the digesters are not operated properly.

Most or all of the above issues can be reduced or eliminated through proper design and selection of equipment. Some suggestions follow.

To solve the above SBR operating problems, Nolasco, Irvine and Manoharan (10) have recommended the following possible solutions:

1. Developing biological SBR operator training programs to complement traditional training with SBR and process control concepts. Developing Water Environment Federation training manuals and video tapes for SBR; offer State SBR Operator Certificates to certify qualified operators.
2. Specifying proper heating, insulation, and operation-and-maintenance procedures to protect exposed equipment from the elements (for example, by using heat tracing and cold-weather greases).
3. Selecting decanters that meet treatment objectives taking into account the effluent quality required, type of processes located downstream of the biological SBR, and available budget.
4. Implementing the use of on-line or portable dissolved oxygen monitors to control blower operation.
5. Taking into account operating conditions when designing pretreatment systems. For example, if the SBR plant is going to be staffed part-time, select self-cleaning pretreatment units.
6. Providing adequate sludge-age control strategies. For example, if sludge-age control is based on mixed liquor concentrations, the supplier or consultant should provide the target concentrations. Automatic solids retention time control could be implemented by using on-line suspended solids monitors.

7. Assessing the impact of aerobic digester operation on phosphorus removal and investigate optimum operating strategies for the biological SBR-aerobic digestion treatment system.

6.5. Online Information

Sequencing batch reactor suppliers, products, news and resources for professionals in the water and wastewater treatment industries can be found from the internet web site, www.wateronline.com (21). For instance, Lemna USA, Inc. is now marketing the Anaerobic Sequencing Batch Reactor (ASBR), which was developed in Minneapolis, MN. and US Filter is demonstrating its Omnipac® SBR Package Plants in Edwardsville, KS. The readers are encouraged to search for more SBR information elsewhere (21–26, 35–37, 46–51).

7. DESIGN EXAMPLES

7.1. Example 1

A plant that has an effluent quality requirement of 30 mg/L BOD_5 and 30 mg/L TSS with no requirement for ammonia or phosphorus removal will have a considerably different design approach than a plant needing to attain 10 mg/L BOD_5, 15 mg/L TSS, 1 mg/L NH_3-N, and 1 mg/L phosphorus.

The influent wastewater characteristics are assumed to be:

225 mg/L	BOD_5
25 mg/L	NH_3-N
225 mg/L	TSS
9 mg/L	phosphorus
5 MGD	design flow
10 MGD	maximum daily flow (2 × design flow)

The effluent requirements are assumed to be:

10 mg/L	BOD_5
15 mg/L	TSS
2 mg/L	NH_3-N
3 mg/L	phosphorus

Design a traditional biological sequencing batch reactor (SBR) system.

Solution:

The effluent requirement dictates nitrification to attain ammonia reduction. Therefore, a low load design with an F:M ratio of 0.05 to 0.1 should be used.

A reasonable value of MLSS concentration should be assumed in the SBR basin at the end of DECANT phase (low water level). The MLSS concentration might be as low as 2000 mg/L and as high as 7500 mg/L. A reasonable design base on the influent characteristics and the effluent quality desired is 4500 mg/L.

At this point, the design of the traditional SBR system departs from the design of an activated sludge system. The MLSS concentration assumed in an SBR design changes continuously throughout the SBR operating cycle, from a maximum at the beginning of the FILL

phase to a minimum at the end of the REACT phase. Other conditions remaining the same, the MLSS at the beginning of the FILL phase will be higher than the corresponding value used in the design of a continuous flow through system.

The MLSS has been selected to be 4500 mg/L at the low water level. With a 4500 mg/L MLSS at the low water level, the MLSS at high water level will be approximately 300 mg/L. Using a conservative design approach, a F:M ratio of 0.064 would be used. The F:M ratios and the MLSS at the low water level establish the reactor volume or size, as shown in the following example:

1. Daily BOD loading (L)
$$= C \times Q$$
$$= 225\,\text{mg/L BOD}_5 \times [8.34(\text{lb/MG})/(\text{mg/L})] \times 5\,\text{MGD}$$
$$= 9383\,\text{lb BOD}_5/\text{d}$$

2. Total MLSS (by weight) in the reactor
$$= (9383\,\text{lb BOD}_5/\text{d})/[(0.064\,\text{lb})(\text{BOD}_5/\text{lb MLSS/d})]$$
$$= 146{,}609\,\text{lb MLSS}$$

3. Reactor volume at low water level
$$= (146{,}609\,\text{lb MLSS})/[(4500\,\text{mg/L})(8.34)]$$
$$= 3.9\,\text{MG reactor volume}$$
$$= (3{,}900{,}000/7.48)\ \text{ft}^3$$
$$= 521{,}390\,\text{ft}^3$$

4. Number of cycles per day
Next select the number of cycles per day; each cycle comprising STATIC FILL, REACT, SETTLE, DECANT, SLUDGE WASTE, and IDLE. This parameter is quite critical in reactor sizing.

   The number of cycles dictates the number of decants per day and, hence, the volume of liquid to be decanted for each cycle. The volume must be selected based on the maximum sustained daily flow. Larger flows would shorten the cycle time and could adversely affect effluent quality if encountered over prolonged periods. Fewer cycles per day produce fewer decants and require a larger volume of variable capacity.

   It has been demonstrated the 5 cycles per day per basin, based on maximum sustained daily flow, is an optimum cycle selection for low load design.

5. Number of basin:
The next step is to determine the number of SBR basins.

   The decision with respect to the number of basins is not unique to an SBR design. A single basin mode of operation can accommodate inflow on a continuous basis, although at least two basins are usually preferred to provide greater flexibility and improved effluent quality.

6. Decant volume:
Next, calculate the volume of liquid per basin per decant.
5 cycles/d = 5 decants per day per basin

   Volume of liquid to be handled per decant (maximum daily sustained flow; for this design, assume 2 times average design flow) = 5 million gallons × 2 = 10 million gallons.

7. Basin design at 10 MGD
   Four basins have been selected for the design example.

$$\text{Volume per decant} = 10{,}000{,}000\,\text{gpd}/(4\text{ basins})/(5\text{ cycles/d/basin})$$
$$= 500{,}000\text{ gallons}$$
$$= \frac{500{,}000}{7.48}\text{ cu. ft.}$$
$$= 66{,}845\,\text{ft}^3$$

8. Reactor size:
   Now calculate the total reactor size.
   Volume per basin at low level (at the end of decant phases)

$$= (521{,}390\,\text{ft}^3)/(4\text{ basins})$$
$$= 130{,}348\,\text{ft}^3/\text{basin}$$

Total volume required per basin $= 130{,}348 + 66{,}845 = 197{,}193\,\text{ft}^3$

The minimum depth after decant is determined as you would determine the depth of a clarifier in a flow through system. Because the SBR process provides for significantly larger surface settling area, a minimum depth of 8 ft. is not uncommon. However, a more conservative design would be a 10 feet minimum depth.

The ratio of minimum depth to maximum depth is determined to be approximately 2/3 in this example. This would mean a basin as shallow as 12 feet would be acceptable. A more conservative design would be to allow for a 10 feet minimum depth after decant. The resulting maximum depth of 15 feet is common and practical from the standpoint of blower discharge pressure.

Maximum depths of 20 feet and more can be used and are particularly suitable when the decanter portion represents a larger fraction of the total volume.

$$\text{Area per basin} = \frac{197{,}193}{15}$$
$$= 13{,}146\text{ sq. ft}$$

Length and width or diameter should be decided in the same manner as in a continuous flow activated sludge plant. It is primarily a function of the physical space available and the type of aeration system used. A unique design characteristics permit the use of square or rectangular, as well as round, tanks (1). Therefore, reduction in concrete using common wall construction can be incorporated.

For this design, assume square basins with common wall construction.

Length = Width = $(13{,}146)^{0.5} = 114.65$ feet

To assure sufficient volume and to simplify design, use 115 feet for the length and width.

Therefore, the final design is as follows:

Number of SBR basins = 4

$$\text{Size of each basin} = (115\text{ ft. length}) \times (115\text{ ft. width}) \times (15\text{ ft. depth})$$
$$= 198{,}375\text{ cu. ft.}$$

Number of cycles per day per basin = 5 each with a cycle time of 4.8 hours

The normal cycle time of 4.8 hours will decrease and the number of cycles per day will increase during storm flows in excess of 10 MGD.

9. Detention time at maximum level

$$= 4(198{,}375\ \text{ft}^3)(7.48\ \text{gal/ft}^3)/[5(1{,}000{,}000\ \text{gpd})]$$

$$= 1.187\ \text{days}$$

$$= 28.49\ \text{hours}$$

10. Detention time at minimum level

$$= 4(130{,}348\ \text{ft}^3)(7.48\ \text{gal/ft}^3)/[5(1{,}000{,}000\ \text{gpd})]$$

$$= 0.78\ \text{days}$$

$$= 18.72\ \text{hours}$$

The desired F:M ratio will decrease with repeating cycles unless the MLSS concentration is adjusted. Therefore, once the MLSS has reached the desired level, sludge wasting at some regular interval, usually the end of each cycle, is necessary.

11. Aeration equipment sizing

Sizing the aeration equipment is done in a similar manner to that used in the continuous flow through an activated sludge system, with one exception. Because the aeration equipment runs for only a portion of the SBR operating cycle (generally a part of FILL plus REACT phases), the calculated daily oxygen requirement must be met in this shorter time period. It will, therefore, increase the size of the aeration equipment accordingly.

Owing to the unique characteristics of the traditional SBR process, the actual oxygen demand for BOD_5 reduction and nitrification may be less than that associated with a flow through activated sludge process.

Some carbonaceous BOD_5 removal probably occurs through anaerobic activity. In addition, extremely low residual oxygen levels are typical during the initial portions of the REACT phase. These conditions lead to an apparent high level of efficiency for the aeration system use.

Nevertheless, prudent design would dictates that a conservative approach be used in the selection of the aeration system.

An experienced SBR engineer sizes the aeration equipment to provide 1.5 lb of oxygen per lb. of BOD_5 applied at a residual dissolved oxygen level of 2.0 mg/L. In addition, if nitrification is required or possible, additional oxygen is provided on the basis of 4.6 lb per lb. of ammonia nitrogen applied.

Although this approach may result in slightly greater aeration capacity, it is important to realize that the aeration cycle will be tailored to use only the actual aeration energy required. It is also necessary to compare competitive systems based on the actual aeration rate provided. Some equipment suppliers use fractional dissolved oxygen residuals and reduced ratios of oxygen to BOD_5 and ammonia nitrogen to justify reduced aeration horsepower. This practice should be recognized as a departure from conservative design standards.

As a typical example, if the cycle time is 4.8 hours based on 5 cycles per day, the aeration time (React time) is 1.58 hours per cycle. At 5 cycles per day per basin and with 4 basins, the aeration time is $1.58 \times 5 \times 4 = 31.6$ hours.

The following is a standard for low load design:

4.8 hours cycle time
1.58 hours of aeration time per cycle
2.23 hours of mixer time per cycle

12. Decanter and associated piping sizing:
During peak flow, the volume of liquid to be decanted per cycle will increase over that volume to be decanted during normal design flow. Therefore, the time allocated for the DECANT phase and the decant rate should be designed to handle the maximum daily flow.

Maximum daily flow per basin = 10 MGD/(4 basins) = 2.5 MGD per basin

Volume of liquid to be decanted per cycle at maximum daily flow

= 2.5 MGD/(5 cycles per day)

= 500,000 gallons per cycle

Decant time = 60 minutes(selected)at maximum daily flow

Decant rate = 500,000 gal/(60 min.)

= 8,333 gal/min

Therefore, the decanter and all downstream piping must be designed to accommodate the above rate of flow form each basin. The decant rate (8333 GPM) is also used for designing the downstream chlorination units.

This example illustrates simplified approach. The recommended design in this chapter has the flexibility to accommodate varying process demands.

7.2. Example 2

The use of microprocessor controlled phases enables the operator to vary the operating strategy of a biological sequencing batch reactor (SBR) process system to suit treatment requirements. Normally, the process follows basic steps of FILL, REACT, SETTLE, and DECANT. The ability to create aerobic or anoxic conditions within the reactor results in flexible operation, better treatment of waste, and optimum effluent quality.

Explain which steps are FILL phases, and which steps area Non-FILL phases.

Solution:

MIXED FILL and REACT FILL are of FILL Phases. REACT, SETTLE, DECANT/ SLUDGE WASTE, and IDLE are of Non-FILL phases.

The various operating phases are further explained in Figure 11.7.

7.3. Example 3

In 1991, the village of Intercourse located in Leacock Township, PA, USA was in need of a new wastewater treatment system. The system was to be efficient, reliable, and simple to operate and maintain. At the time SBR technology was emerging as an efficient and economical alternative to conventional flow-through activated sludge system. Knowing this, the township's engineer turned to the manufacturer of a biological sequencing batch reactor (SBR) process system and began looking at using sequencing batch reactor technology as a possible system for Leacock Township Sewage Treatment Plant.

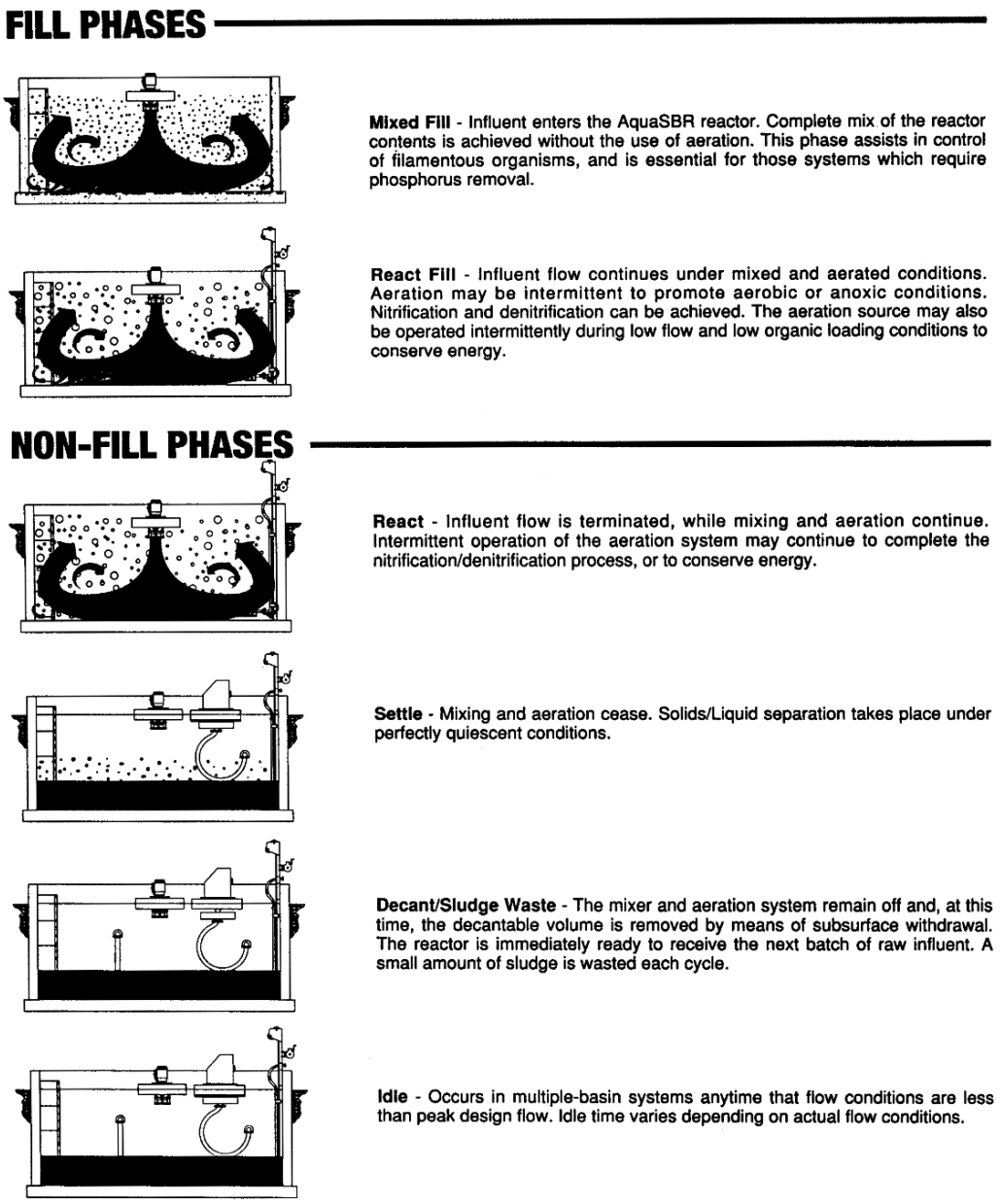

Fig. 11.7. Fill phases and nonfill phases of biological SBR (4).

The sewage flows during the design phase were:

1. Design daily flow = 0.30 MGD (1135 m^3/day)
2. Peak flow = 0.75 MGD (2839 m^3/day)

What was the engineering solution?

Solution:

Following the evaluation of other treatment methods, the Township's engineer and the SBR manufacturer's representative proposed to the township that the best option was a state-of-the-art biological SBR. The new plant would consist of a dual-basin system that would be designed to treat an average flow of 0.30 MGD and a peak flow of 0.75 MGD. Each of the basins would include retrievable coarse-bubble diffusers, mixers and decanters.

In September of 1991, the new SBR System went on-line and for the past many years has consistently met the plant's treatment objectives of BOD_5, TSS, NH_3-N and phosphorus. Over the course of several years, the organic and ammonia loadings to the plant increased to the point where they regularly exceeded the nominal design values. The plant was originally designed to handle influent BOD_5 of 250 mg/L but was receiving an average of 340 mg/L by proper management of the existing aeration capacity and careful solids management, the plant operator was able to maintain effective levels of treatment, however, it became apparent that the long-term solution to the excessive loading was to provide additional oxygen.

Engineers evaluated several possible solutions for the Leacock plant and presented recommendations to the Township's engineer.

One option was to build a third SBR basin, but obviously this would be costly. A second option was to increase the amount of air being put into the existing SBR reactors. This would involve replacing the existing retrievable coarse-bubble diffused air system with a retrievable fine-bubble diffusers system. Doing so would increase the amount of oxygen transfer thus promoting successful BOD reduction and nitrification.

To minimize cost as well as maintain both SBR basins in operation during the aeration system upgrade, the SBR manufacturer proposed that the change-out be done using the same track beams for the coarse bubble diffusers racks. The new fine-bubble assemblies would simply be retrofitted onto the existing tracks.

In June of 2000, the aeration expansion at Leacock Township was completed in just one day without dewatering the basins. Both reactors, each with 2 (10 tubes) coarse-bubble diffuser racks were retrofitted with (20 tubes) fine bubble assemblies. The existing blowers were resheaved to operate at a higher capacity and the motors upgraded from 15 to 20 HP, so there was no need to purchase new blowers.

The aeration upgrade has saved Leacock Township considerable expense compared to building a third reactor. The upgrade has allowed the SBR to be rerated by the State government's Department of Environmental Protection to a higher organic and ammonia loading capacity, and extended the life of the facility well into the foreseeable future.

Recent performance has been excellent, with effluent BOD_5 less than 5 mg/L, TSS less than 10 mg/L, ammonia nitrogen less than 0.7 mg/L, and total phosphorus less than 2 mg/L.

This engineering solution was sound and cost-effective because of the following factors:

1. All components are retrievable and accessible
2. Tolerates variable hydraulic loads
3. Controls filamentous growth
4. Flexible in handling varying influent and effluent conditions
5. Provides quiescent settling
6. Provides separation of aeration and mixing
7. Lower installation costs

Simple to expand or upgrade.

7.4. Example 4

Foxwoods Casino, owned and operated by the Mashantucket Pequot Tribe, opened in February of 1992 with a compact 0.2 MGD wastewater treatment system. The success of the casino and resort far exceeded expectations, and the amount of domestic waste necessitated the immediate upgrade of their wastewater treatment system.

Several options for expending the system were examined including: rotating biological contractors (RBC), oxidation ditches, a conventional activated sludge system, and biological sequencing batch reactors (SBR).

The following were the given information:

1. Plant name: Mashantucket Pequot WWTP (Foxwoods Casino), USA
2. Type of plant: municipal/resort
3. Design daily flow = 2.6 MGD (9,842 m^3/day)
4. Peak flow = 3.14 MGD (11,886 m^3/day)

What was the final engineering solution to this case?

Solution:

It was determined that a dual-basin biological SBR system would best meet the needs of the project.

The SBR was selected because of its ability to handle high strength, peak hydraulic and organic loading, while consistently providing high quality effluent mandated by the Mashantucket Pequot Tribe Council. In addition, installing the SBR system would allow the Casino's wastewater treatment system to be expanded with minimal cost or disruption.

The new 1.0 MGD SBR treatment system went on-line in July of 1993 and had the flexibility to meet future needs as the casino and the resort area expanded. At that time, the casino was treating an average of 0.4 MGD.

The casino and resort continued to grow at a fast pace and once again the wastewater treatment system had to be expanded. In 1998, a new upgraded system would treat an average flow of 2.6 MGD. This upgrade was accomplished by installing a third SBR basin and two 6-disk Disk Filter units.

A future SBR #4 basin has already been constructed for future growth. It is placed adjacent to the SBR #3 basin.

The SBR systems installed there operate on a simple concept of introducing a quantity of waste to a reactor, treating the waste in an adequate time period, and subsequently discharging a volume of effluent plus waste sludge that is equal to the original volume of waste introduced

Table 11.4
Average SBR performance data July 1998–October 2000 at Mashantucket Pequot WWTP (Foxwoods Casino), USA (5)

	Design influent	Current influent	Current effluent	Permit effluent
Avg. flow, MGD	2.60	1.12	—	—
Peak flow, MGD	3.14	2.26	—	—
BOD_5, mg/L	434	249	1.5	5
TSS, mg/L	709	414	2.9	5
TKN, mg/L	118	50	3.2	9
Total P, mg/L	17	14	1.0	3

to the reactor. This "Fill and Draw" principle of operation involves the basic steps of Fill, React, Settle, Decant, and Sludge Waste. The system has been designed to include seven individual phases of operation both inclusion or duration of any individual phase is based upon specific waste characteristics and effluent objectives.

Where and when nutrient removal is to be required, a simple adjustment to the SBR's operating strategies permits nitrification, denitrification and biological phosphorus removal. Optimum performance is attained when two or more reactors are used in a predetermined sequence of operation.

The plant's current average design daily flow is 2.60 MGD (9842 m^3/d) and the peak design daily flow is 3.14 MGD (11,886 m^3/d).

Table 11.4 indicates excellent average operating data for the period of July 1998 to October 2000.

The Mashantucket Pequot Tribe chose SBR system for Foxwoods Casino because it can handle strong peak hydraulic and organic loadings although providing high quality effluent, and it provides expansion with minimal cost and disruption. All components are retrievable and accessible.

7.5. Example 5

Millions of people visit Biloxi, Mississippi, USA for it casinos, beaches, golf courses, and other attractions. Some even visit the city to tour the Keegan Bayou Wastewater Treatment Plant.

The wastewater treatment facility at Biloxi was originally constructed with primary clarifiers and trickling filters in the late 1940s. It was then expanded in the 1960s, the early 1980s, and again in March of 1998. The most recent expansion was prompted by the city's experience of explosive development in the east Biloxi area and by the plant being subject to more stringent effluent regulations. Keegan Bayou WWTP remained in full operation during this latest expansions, which was also completed ahead of schedule and under budget.

The following were the design conditions and information:

1. Plant name: Keegan Bayou WWTP, USA
2. Type of plant: municipal/ domestic
3. Design daily flow: average flow of 10 MGD (37,854 m^3/d)
4. Peak Flow = 17 MGD (64,352 m^3/d)

Table 11.5
Average SBR performance data in 1998 at Keegan Bayou WWTP, USA

Loading	Design influent (4-basin)	Current influent* (2-basin)	Current effluent* (2-basin)	Bay permit effluent**
Ave flow, MGD	10	4.5	—	—
Peak flow, MGD	17	8.5	—	—
BOD_5, mg/L	140	252	6	30
TSS, mg/L	156	267	5	30
NH_3-N, mg/L	40	15	1	15

* Only two of the four basins are currently needed for operation.

** The plant was designed to discharge to Keegan Bayou and meet effluent objectives of 10 mg/L BOD_5, 30 mg/L TSS and 2 mg/L NH_3-N. The plant chooses to discharge its effluent to the back Bay of Biloxi and is required to meet the effluent objectives shown in the table above.

What was the engineering solution to accommodate Keegan Bayou's small footprint requirement?

Solution:

To accommodate the growing demands on Keegan Bayou's wastewater treatment system, the recent expansion included replacing the conventional secondary treatment process with a state-of-the-art, four-basin biological SBR system. A SBR system was chosen because of the small footprint it required. Its design also enabled modification and continued use of the facility's existing components. The old trickling filters are now used as equalization basins and the old primary clarifiers are now as sludge thickening tanks. Construction cost for this expansion was approximately $12 million US dollars.

The plant's new biological SBR treatment process nearly tripled its capacity, from 3.5 MGD to 10 MGD. Keegan Bayou's increased capacity allows it to now serve all of east Biloxi along with an area of casino-hotel chains referred to as "Casino Row". The SBR system will also accommodate the city's predicted future development demands.

The SBR system installed there also operate on a simple concept of introducing a quantity of waste to reactor, treating the waste in an adequate time period, and subsequently discharging a volume of effluent. Sludge is also wasted to maintain biomass concentration. This "Fill and Draw" principle of operation involves the basic steps of Fill, React, Settle, Decant, and Sludge Waste. The system has been designed to include seven individual phases of operation. The inclusion or duration of any individual phase has been based upon specific waste characteristics and effluent objectives.

Because nutrient removal is required, a simple adjustment has been made to the SBR's operating strategies permits nitrification, denitrification, and biological phosphorus removal (Table 11.5)

Storm flow management was an important consideration in the design of Keegan Bayou's biological SBR system, because it needs to retain sufficient biomass for subsequent treatments during peak wet weather events.

In fact, Keegan Bayou has already successfully managed peak flow periods in excess of its design flow. This occurred during Hurricane George, in 1998. Peak flows during the hurricane

surpassed 14 MGD, with only two of the four basins in operation. Also, effluent BOD_5 and TSS levels never exceeded permit limits during the storm (average effluent BOD_5 was 4.0 mg/L and average effluent TSS was 8.6 mg/L).

Predicted future development in the east Biloxi area was also an important factor in the biological SBR system design chosen for the Keegan Bayou WWTP. The design incorporates two additional basins to accommodate future growth demands. Table 11.5 summarizes the excellent average operating data in 1998. It is important to note that only two of the four basins were needed for operation in 1998. The plant was designed to meet effluent objectives of 10 mg/L BOD_5, 30 mg/L TSS and 2 mg/L NH_3-N. The plant chooses to discharge its effluent to the back Bay of Biloxi and is required to meet the effluent objectives shown in the table in this chapter.

7.6. Example 6

A wastewater plant, which represents the operational flexibility of the biological SBR, is near Lake Sanitary, Clear Lake, Iowa, USA. Wesley M. Shubert designed the SBR Process, while Rust Engineering did the overall plant design.

1. The effluent requirements by the Iowa Department of Natural Resources to allow a discharge permit are
   a. CBOD = 10 mg/L
   b. TSS = 10 mg/L
   c. NH_3-N = 2 mg/L
2. Design parameters were:
   a. at average present flow of 2.7 MGD = 2297 CBOD lb/d
   b. at average daily flow of 3.4 MGD = 4197 CBOD lb/d
   c. at max monthly flow of 5.7 MGD = 5419 CBOD lb/d
   d. at max daily flow of 8.2 MGD = 6269 CBOD lb/d

Please visit the SBR biological plant at Clear Lake, Iowa. Find out how the SBR plant was designed, and how it performed in the second half of 1999. Comment on its operational performance in comparison with the Iowa effluent requirements.

Solution:

1. Design was based on F/M ratio of 0.52, MLSS of 4,500 mg/L and five cycles per day per basin. Four basins were used each 91 ft. diameter with a high water level of 19.5 ft., and a decant flow rate of 5,674 gallons per minute.
2. Operational performance in the period of May–December 1999 is documented in Table 11.6.
3. Comments by a licensed Professional Engineer:
   (a) At Clear Lake, SBR is an excellent alternative to a conventional activated sludge process system.
   (b) The SBR plant is capable of achieving high reduction of CBOD, TSS and ammonia-nitrogen, meeting the plants effluent requirements (CBOD = 10 mg/L; TSS = 10 mg/L; ammonia-nitrogen = 2 mg/L) at all times.
   (c) The SBR plant has successfully accomplished carbonaceous oxidation, nitrification and denitrification, and is very suitable to small community, such as Clear Lake, Iowa.
   (d) The influent wastewater flow varied significantly between 1.5 and 9.9 MGD. SBR was very flexible in handling such a large flow change.

Table 11.6
Operational performance of biological SBR plant at Clear Lake, Iowa, USA

	Influent				Effluent		
Operational Period	Flow MGD	CBOD lb/day	TSS lb/day	NH_3-N mg/L	CBOD mg/L	TSS mg/L	NH_3-N mg/L
(a) May 1999							
Month average	4.9	2235	4223	4.78	3.1	7.8	0.20
Maximum day	9.9						
Minimum day	2.7						
(b) June 1999							
Month average	3.5	2069	3297	7.85	3.3	3.8	0.14
Maximum day	5.2						
Minimum day	2.5						
(c) July 1999							
Month average	3.6	2545	4336	9.07	3.2	4.8	0.01
Maximum day	8.6						
Minimum day	2.2						
(d) December 1999							
Month average	1.6	1645	2413	15.2	2.94	4.63	0.06
Maximum day	1.7						
Minimum day	1.5						

(e) SBR offers additional features to Clear Lake, Iowa. These features include easy installation, simple operation, lower maintenance than most activated sludge variation, and energy efficiency.

7.7. Example 7

Figure 11.9 shows a typical biological operating sequence (52, 53–55). Explain the purpose of each operating step. Discuss the operation and maintenance (O and M) requirements. Discuss the most popular methods for residue disposal at a biological SBR plant.

Solution:

1. Purpose of Each Operating Step:
   (a) The SBR system shown in Figure 11.9 is one of common biological sequencing batch reactors, which are a variation of the conventional activated sludge treatment system in which equalization, aeration, clarification, and sludge wasting processes are carried out sequentially in the same tank. SBRs consist of a single tank equipped with an inlet for raw wastewater, air diffusers with associated blowers and piping for aeration, a sludge draw-off mechanism at the bottom to waste sludge, a decant mechanism to remove supernatant after settling, and a control mechanism to time and sequence processes. SBRs operate in cycles of five periods carried out in sequences as follows: FILL, REACT (aeration), SETTLE (clarification), DRAW (decant), and IDLE (sludge wasting). These processes are controlled by time to achieve the objectives of operation. They are discussed further in the following paragraphs. Figure 11.9 shows a typical single cycle.

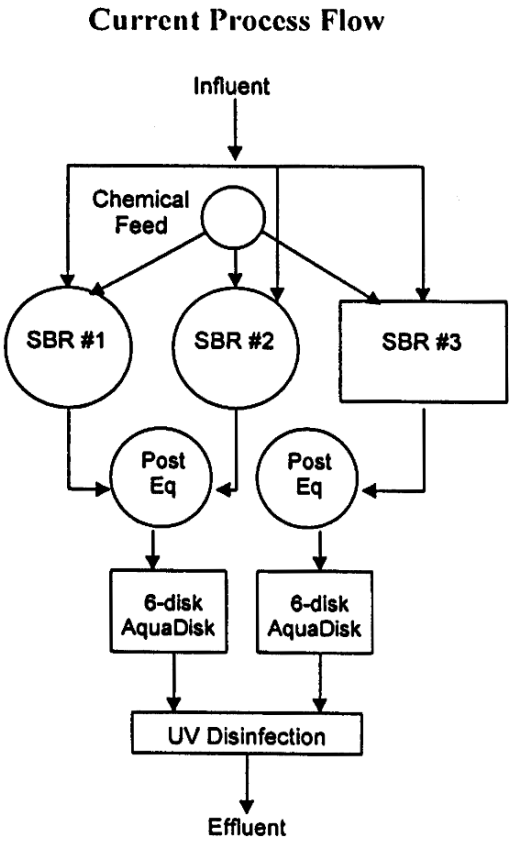

Fig. 11.8. Process flow diagram of Mashantucket Pequot wastewater treatment plant (Foxwoods Casino) (5).

(b) FILL
The purpose of the FILL operation is to add raw wastewater to the reactor. During the FILL phase, performance standards may require alternating conditions of low and high dissolved oxygen (DO) concentrations. Periods of aeration during FILL are critical to the development of organisms with good settling characteristics. Conversely, period of zero DO (anaerobic conditions) or low DO (anoxic conditions) are necessary for biological nutrient removal of nitrogen and phosphorus.

(c) REACT
The purpose of the REACT phase is to complete the reactions initiated during the FILL stage. Depending on design type, influent flow may be diverted to another reactor during this phase and aeration continues during this period separated by long distances, baffles, etc. In other designs, organic removal occurs during this stage. Nitrification (ammonia removal) may also occur during this phase if loading is low enough compared to Mixed Liquor Suspended Solids (MLSS) (i.e., high Solids Retention Time, SRT).

(d) SETTLE
The purpose of the SETTLE phase is to allow solids separation to occur in the system while providing a clarified supernatant to be discharged as effluent. In the SETTLE mode,

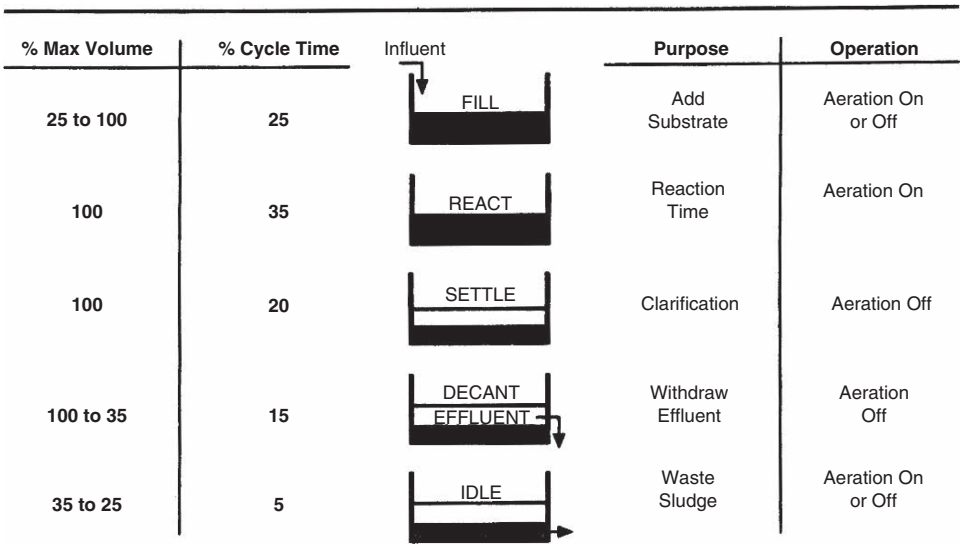

Fig. 11.9. Typical biological SBR operating sequence (52, 53).

reactor contents are completely quiescent, eliminating the short–circuiting of continuous flow clarifiers.

(e) DRAW (or DECANT)
The purpose of the DRAW phase is to remove the clarified supernatant from the reactor as final effluent. Floating and adjustable weirs are the most popular decanting mechanisms for this phase of treatment, but submersible pumps are also used.

(f) IDLE
The purpose of the IDLE phase is to provide time for one reactor to complete its fill cycle before switching to another unit. IDLE is not a necessary phase and can be eliminated. Depending upon the process and treatment goals, aeration, mixing, or sludge wasting can occur during the IDLE phase.

Continuous influent types of SBRs do not have an IDLE phase.

2. Operation and Maintenance Requirements
O and M requirements for SBRs are minimum compared to other conventional activated sludge treatment systems. However, SBR plants should be checked daily by personnel experienced in operating biological treatment systems. Depending on the size of the facility and complexity of treatment processes (i.e., nitrogen and phosphorus removal), the operator should be present from 2 to 8 hours per day. Adequate time should be allocated for process control, sampling, O and M. and daily record keeping. Unless these plants receive at least some daily attention and maintenance from a qualified operator, effluent quality will eventually become unsatisfactory. To operate the treatment system properly, the following operating parameters should be monitored at least weekly:
(a) 30 to 60 minute sludge settling volume;
(b) MLSS concentration;
(c) DO concentration; and
(d) Decant heights

These parameters should be checked against predetermined target values to evaluate the performance of the system. Typical operating target ranges for SBRs are as follows:
(a) SRT = 20 to 30 days;
(b) MLSS = 2000 to 6000 mg/L;
(c) F/M = 0.08 to 0.16; and
(d) pH = 7.0 to 8.0

Regular preventive maintenance (PM) is required to keep the equipment in optimal condition. A formalized PM program should be established based upon the recommendations of the equipment manufacturers. This program should include a listing of all equipment, required PM tasks, and the frequency with which the tasks should be performed. Typical equipment that requires routine PM includes pumps, blowers, air diffusers, and automatic controllers.

3. Residue Generation and Disposal
Generally, SBRs generate the same quantities of sludge as extended aeration activated sludge facilities. Excess or waste activated sludge may be typically aerobically digested, dewatered on drying beds, and applied to the land.

7.8. Example 8

Figure 11.6 shows the unit construction costs (1998 US Dollars) of biological SBR as a function of plant capacity. Please explain how the 1998 cost can be updated to the present or future cost.

Solution

The following equation can be used for converting the past cost to the present or future cost, or vise versa.

$$\text{Cost}_b = \text{Cost}_a(\text{Index}_b)/(\text{Index}_a)$$

where
Cost_a = the cost in the month-year of a, \$
Cost_b = the cost in the month-year of b, \$
Index_a = the Cost Index in the month-year of a
Index_b = the Cost Index in the month-year of b

The following two kinds of Cost Indices are frequently used for cost conversion or updating:

1. ENR Cost Indices (57)
2. CE Cost Indices (56)

where
ENR = Engineering News Records
CE = Chemical Engineering

In general, for cost estimation of a process equipment itself, CE Cost Indices are used, for cost estimation of a process system, ENR Cost Indices are used (55–59). Wang, Aulenbach and Wang (58) discuss the ENR Cost Indices and applications.

7.9. Example 9

Tanning is a chemical process that converts raw hides and skins into a stable material. Vegetable, mineral and other tanning agents may be used – either separately or in combination – to produce leather with different qualities and properties. Trivalent chromium is the main tanning agent because it produces a modern, thin, light leather suitable for shoe uppers, clothing and upholstery. However, recent limits for discharge to the environment have limited trivalent chromium discharge to levels as low as 2 mg/L in wastewaters

The United Nations (38) has promoted a physical-chemical sequencing batch reactor (PC-SBR) technology for recovery of trivalent chromium from the spent tannery liquors and its reuse. Please describe the operating sequence and the purposes of each operating step of PC-SBR. Explain the chemistry of the trivalent chromium recovery process. Investigate a real case history at the Germanakos SA Tannery near Athens, Greece, and discuss the economic benefit of this new technology. Make your constructive recommendations on the newly developed PC-SBR process system.

Solution:

1. The detailed theory and principals of a PC-SBR process system can be found elsewhere (2, 3). The operating sequence of a PC-SBR process is very similar to that of a typical biological SBR process system shown in Figure 11.9, except that (a) chemical are used for physical-chemical reactions (instead of using microorganisms for biochemical reactions; and (b) mechanical mixing are used (instead of aerating the bioreactor for both mixing microorganisms and supplying oxygen). Specifically the PC-SBR process system used at the Germanakos SA Tannery near Athens, Greece, consists of the following operating steps in sequence:
   (a) FILL (step 1)
   The purpose of the FILL operation is to add the spent tanning liquor to the reactor. No mixing is required.
   (b) REACT-1 (step 2)
   The purposes of the REACT-1 phase are: (a) to add chemicals (magnesium oxide and polyelectrolyte) to the reactor; and (b) to mix the mixture of the spent tanning liquor and chemicals by mechanical means (although aeration can also be applied for mixing). Soluble trivalent chromium ions are converted to chromium hydroxide during this step, forming insoluble chromium hydroxide flocs.
   (c) SETTLE (step 3)
   The purpose of the SETTLE phase is to allow solids separation to occur by gravity in the system while providing a clarified supernatant to be discharged as effluent. In the SETTLE mode, reactor contents are completely quiescent, eliminating the short-circuiting of continuous flow clarifiers. The insoluble chromium hydroxide flocs settle at the bottom of the reactor.
   (d) DRAW/DECANT (step 4)
   The purpose of the DRAW (or DECANT) phase is to remove the clarified supernatant from the reactor as final effluent. Floating and adjustable weirs are the most popular decanting mechanisms for this phase of operation, although submersible pumps are also used. After this phase is over, only the insoluble chromium hydroxide solids remain at the bottom layer of the reactor.
   (e) REACT–2 (step-5)
   The purpose of the REACT-2 phase are: (a) to add sulfuric acid to convert insoluble chromium hydroxide to soluble basic chromium sulfate during mixing; and (b) to discharge

and reuse the recovered basic chromium sulfate solution in the tanning process. At the end of the REACT-2 phase, the reactor is empty.

(f) IDLE (step-6)
The purpose of the IDLE phase is to provide time for one reactor to complete its fill cycle before switching to another PC-SBR reactor unit. IDLE is not a necessary phase and can be eliminated.

2. Continuous influent types of PC-SBRs do not need an IDLE phase
3. The chemistry of the process is presented below:
   (a) Chemical Precipitation

$$MgO + H_2O \rightarrow Mg(OH)_2$$

$$2\,Cr^{3+} + 3\,Mg(OH)_2 \rightarrow 2\,Cr(OH)_3(\text{small insoluble flocs}) + 3\,Mg^{2+}$$

In the chemical precipitation reaction, soluble trivalent chromium ions are converted to insoluble, settable chromium hydroxide solids.

   (b) Flocculation

$$Cr(OH)_3 + \text{polyelectrolytes} \rightarrow Cr(OH)_3 \text{ complex (big insoluble flocs)}$$

In the flocculation process, small insoluble chromium hydroxide solids agglomerate with polyelectrolytes forming visible, big chromium hydroxide complexes.

   (c) Acidification

$$2\,Cr(OH)_3 \text{ complex} + 2\,H_2SO_4 \rightarrow 2\,Cr(OH)(SO_4) + 4\,H_2O$$

In the acidification reaction, the insoluble chromium hydroxide complexes are solubilized by sulfuric acid, forming soluble basic chromium sulfate, which can be, in turn, reused in the subsequent tanning process for cost saving.

4. Real Case History and Financial Benefits

The Germanakos SA tannery near Athens in Greece was founded in 1976. Today it recycles its chromium for reuse, and still produces good quality upper leather from cattle hides (8). Processing 2200 tons per year and with an annual turnover of US$ 8.4 millions (1995 dollars) and a staff of 65.

At the tannery, tanning of hides is carried out with basic chromium sulfate, $Cr(OH)(SO_4)$, at a pH of 3.5 to 4.0. After tanning, the solution is discharged by gravity to a collection pit. The liquor is sieved during this transfer to remove particles and fibers that have come from the hides. The liquor is then pumped to the treatment tank and a calculated quantity of magnesium oxides is added with stirring until the pH reaches at least 8. The stirrer is switched off and the chromium precipitates as a compact sludge of $Cr(OH)_3$. After settling, the clear liquid is decanted off. The remaining sludge is dissolved by adding a calculated quantity of concentrated sulfuric acid (H_2SO_4) until a pH of 2.5 is reached. The liquor now contains $Cr(OH)(SO_4)$ and is pumped back to a storage tank for reuse.

In conventional chrome tanning processes 20% to 40% of the chrome used is discharged into wastewater. In the new process 95% to 98% of the waste trivalent chromium ions can be recycled.

Adoption of the PC-SBR process systems for wastewater treatment and chromium recycle is a typical cleaner production technology (38), which has many advantages: (a) very little changes to production process; (b) more consistent product quality; (c) easier to monitor

amounts of water and process chemical used; and (d) much reduced chromium content in effluent water.

The economic benefits are significant. For the Germanakos tannery, which has a chrome recycling capacity 12 m^3/d, the approximated costs were as follows:

(a) savings = US$ 73,750 per year
(b) operating cost = US$ 30,200 per year
(c) total net savings = US$ 43,550 per year
(d) capital investment = US$ 40,000
(e) pay back = 11 months

Saving can be made with any plant processing more than 1.7 m^3/d.

Recommendations: (a). It is recommended that other tanneries also adopt the PC-SBR chromium recovery process, and (b) multiple PC-SBR system may also be adopted.

7.10. Example 10

Propose a 2-PC-SBR (physicochemical sequencing batch reactor) process system for treating the same spent tannery liquor at Germanakos SA Tannery near Athens, Greece, aiming at chromium recycle. Assume the net settled sludge (chromium hydroxide) production flow rate is estimated to be 10% of spent tannery liquor's influent flow rate.

Solution

A 2-PC-SBR process system will consist of the following:

1. A large PC-SBR is to be sized and designed according to the influent flow of the spent tannery liquor, and with the process steps/sequence of:
   (a) FILL (step 1)
   (b) REACT-1 (or simply REACT; step 2)
   (c) SETTLE (step 3)
   (d) DRAW/DECANT (step 4)
   (e) IDLE (step 5) -optional
   The above steps 1 to 5 are identical to that of steps 1 to 5 in Example 9, except that after DRAW/DECANT phase (step 4) is over, the insoluble chromium hydroxide solids are discharged to a small PC-SBR specified in below:
2. A small PC-SBR is to be sized and designed according to the small sludge flow rate (containing chromium hydroxide solids) from the DRAW/DECANT phase (step 4) of the aforementioned large PC-SBR. The volume of the small PC-SBR is about 10% of the large PC-SBR because the small sludge flow rate (containing chromium hydroxide solids) has been estimated to be 10% of spent tannery liquors influent flow rate to the large PC-SBR. The small PC-SBR will receive the small sludge flow for acidification treatment, and will have the following operating steps:
   (a) FILL (step 1)
   (b) REACT-2 (or simply REACT, step 2)
   (c) SETTLE (step 3) – optional
   (d) DRAW/DECANT (step 4) – optional
   (e) IDLE (step 5) – optional
   The common operating phases (step 1, 3, 4 and 5) are self-explanatory. REACT –2 here is similar to REACT-2 in Example 9.

3. Conclusion: The 2 PC-SBR process system is more cost-effective from both operating cost-saving and capital cost-saving viewpoints. The large PC-SBR can be constructed using regular inexpensive material, because no acidification reaction will be involved. The small PC-SBR will be used for acidification reaction, therefore, it must be constructed using noncorrosive, expensive material, such as noncorrosive stainless steel or ceramic material.

In previous engineering case history (Example 9), only one large PC-SBR is used for all physicochemical reactions (including acidification). Entire large PC-SBR reactor unit must be constructed using noncorrosive, expensive materials.

7.11. Example 11

Discuss (a) the typical cycle for a single tank in a dual tank SBR system designed for biological nitrification; and (b) the typical nitrification performance according to historical information.

Solution:

Each supplier of traditional biological SBR system equipment has their own approach to design. Some SBR systems are custom designed and the uniqueness of each of these systems reflects the preferences of the design engineer. Designs include the use of different tank configurations, different system hydraulics and a variety of options for aeration, mixing, effluent discharge, and sludge wasting. Systems are normally configured to vary their operation automatically in response to changes in influent flow rate, or to allow the operator to initiate changes to the total cycle time or individual step times, or to make changes during each step (e.g., change length of time for aeration or mixing during fill step). The steps and associated conditions and purpose of a complete, typical cycle for a single tank operated as part of a SBR system designed to achieve nitrification are described in Table 11.7. Nitrification takes place during the react phase and during the portions of the fill period when aeration is practiced.

To design SBRs for nitrification, an adaptation of the approach used in the design of complete mix systems is normally acceptable. The specific calculation procedure will be dictated by the characteristics of the selected SBR system. The most important calculation steps are to determine the minimum required aerobic solids residence time, and to determine the minimum volume requirements that will assure adequate time for settling and decanting. Other critical parameters for the design of the SBR system can be determined from information presented elsewhere (49).

SBR systems are typically designed and operated at long solids residence times ($>$ 15 days) and low F/M (less than 0.1 kg BOD_5/kg MLSS/d). Consequently, partial or complete nitrification is nearly always observed (49, 50). In a recent evaluation of 19 SBR treatment plants (50, 51) (all originally designed for nitrification), influent and effluent ammonia-nitrogen data were reported for eight of the plants (Table 11.8). The average effluent ammonium-nitrogen concentration for the eight plants was less than 2.0 mg/L, implying that a high degree of nitrification was achieved in all cases. These efficiencies reflect the long design solids residence times that are employed and operations that are generally well below the design flow.

Table 11.7
Typical cycle for a single tank in a dual tank SBR system designed for nitrification

Step	Conditions	Purpose
FILL	Influent flow into SBR Aeration occurs continually or intermittently Time = half of cycle time	Addition of raw wastewater to the SBR; COD removal and nitrification
REACT	No influent flow to SBR Aeration Time typically = 1 to 2 hours (varies widely depending on nitrification kinetics, waste strength, and amount of aeration during fill)	Carbonaceous oxidation and nitrification
SETTLE	No influent flow to SBR No aeration Time = approximately 1 hour (depends on settling characteristics)	Allow SS to settle, yielding a clear supernatant
DRAW	No influent flow to SBR No aeration Effluent is decanted Time = 1 hour (variable)	Decant – remove clarified effluent from reactor, 15 to 25 percent of the reactor volume is typically decanted, depending on hydraulic considerations and SBR manufacturer's design
IDLE	No influent flow to SBR No aeration Sludge is wasted Time = variable (determined by flow rate)	Multi-tank system, which allows time for one reactor to complete the fill step before another starts a new cycle; waste sludge – remove excess solids from reactors

Source: US EPA.
Note: A typical total cycle time is 4 to 6 hours.

Table 11.8
Nitrification performance information for SBR operating plants in the USA

Plant Location	Period of Evaluation	Wastewater Flow		Percent of Design Flow	BOD_5, mg/L		Ammonia-N, mg/L	
		m^3/d	MGD		Influent	Effluent	Influent	Effluent
Buckingham, PA	04/89–04/91	439	0.116	49	324	8	25.3	1.1
Clarkston, MI (Chateau Estates)	11/89–04/91	208	0.055	50	192	12	39.1	1.7
Grundy Center, IA	12/89–11/90	2,176	0.575	72	195	4	15.8	1.2
Marlette, MI	07/90–06/91	1,578	0.417	60	103	4	10.1	0.5
Mifflinburg, PA	10/88–03/91	2,763	0.73	81	105	12	7.8	0.4
Monticello, IN (White Oaks Resort)	10/89–05/91	15	0.004	8	131	5	3.1	0.3
Muskegon Heights, MI (Clover Estates)	01/88–10/90	132	0.035	78	185	9	21.2	0.7
Windgap, PA	02/90–10/90	2,116	0.559	56	160	7	12.9	0.6

Source: US EPA.
* Average monthly values based on all data available.

NOMENCLATURE

BOD_5 = 5-day biochemical oxygen demand
C = concentration, lb/MG
CE = chemical engineering
COD = chemical oxygen demand
$Cost_a$ = the cost in the month-year of a, $
$Cost_b$ = the cost in the month-year of b, $
DAF = dissolved air flotation
ENR = Engineering News Records
F:*M* = food to microorganisms ratio
gpd = gallons per day
$Index_a$ = the Cost Index in the month-year of a
$Index_b$ = the Cost Index in the month-year of b
L = BOD_5 loading, lb BOD_5/day
LNAPL = light nonaqueous phase liquid
MG = million gallons
MGD = million gallons per day
MLSS = mixed liquor suspended solids
NH_3-N = ammonia nitrogen
PC = physicochemical
PC-SBR = physicochemical sequencing batch reactors
Q = flow, MGD, or MG/day
SBR = sequencing batch reactor
TSS = total suspended solids

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12
Oxidation Ditch

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract An oxidation ditch is a modified activated sludge biological treatment process that uses long solids retention times (SRTs) to remove biodegradable organics. The typical oxidation ditch is equipped with aeration rotors or brushes that provide aeration and circulation. The wastewater moves through the ditch at 1 to 2 ft/s. The ditch may be designed for continuous or intermittent operation. Because of this feature, this process may be adaptable to the fluctuations in flows and loadings associated with recreation area wastewater production. Several manufacturers have developed modifications to the oxidation ditch design to remove nutrients in conditions cycled or phased between the anoxic and aerobic states.

This chapter covers all aspects of the process including process description, applicability, design criteria, performance, package oxidation ditch plants, operation and maintenance, design parameters and procedure, costs and a worked out design example.

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Key Words Oxidation ditch • wastewater treatment • rotors • BOD • nutrients removal • design procedure • costs.

1. INTRODUCTION

The oxidation ditch, developed in the Netherlands, is a variation of the extended aeration process that has been used in small towns, isolated communities, and institutions in Europe and the United States. The typical oxidation ditch (Figure 12.1) is equipped with aeration rotors or brushes that provide aeration and circulation. The wastewater moves through the ditch at 1 to 2 ft/s. The ditch may be designed for continuous or intermittent operation. Because of this feature, this process may be adaptable to the fluctuations in flows and loadings associated with recreation area wastewater production (1).

2. PROCESS DESCRIPTION

An oxidation ditch is a modified activated sludge biological treatment process that uses long solids retention times (SRTs) to remove biodegradable organics. Oxidation ditches are typically complete mix systems, but they can be modified to approach plug flow conditions. Typical oxidation ditch treatment systems consist of a single or multichannel configuration within a ring or oval basin. As a result, oxidation ditches are called "racetrack type" reactors (2). Horizontally or vertically mounted aerators provide circulation, oxygen transfer, and aeration in the ditch. The cross-sectional area of the ditch is commonly 4 ft to 6 ft deep, with 45° sloping sidewalls. Oxidation ditch systems with depths of 10 ft or more with vertical sidewalls and vertical shaft aerators may also be used. Ditches may be constructed of various materials, including concrete, gunite, asphalt, or impervious membranes. Concrete is the most common. L- and horseshoe-shaped configurations have been constructed to maximize land usage (3).

Preliminary treatment, such as bar screens and grit removal, normally precedes the oxidation ditch. Primary settling before an oxidation ditch is sometimes practiced, but is not typical in this design. Tertiary filters may be required after clarification, depending on the effluent

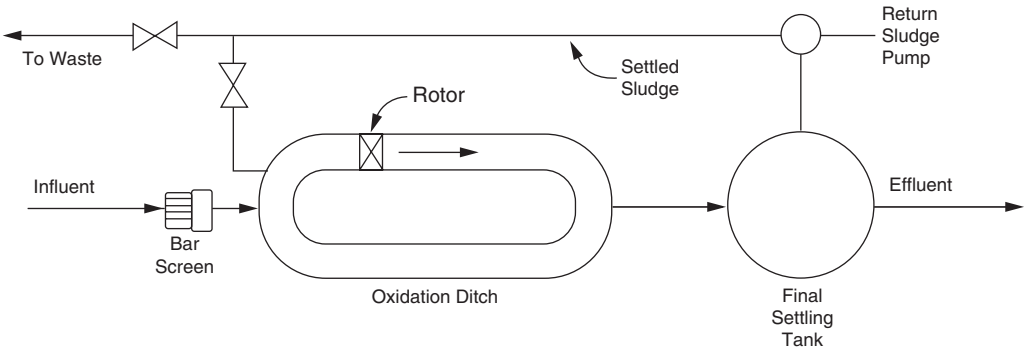

Fig. 12.1. Typical oxidation ditch flow diagram (1).

requirements. Disinfection is required and reaeration may be necessary before final discharge. Flow to the oxidation ditch is aerated and mixed with return sludge from a secondary clarifier. A typical process flow diagram for an activated sludge plant using an oxidation ditch is shown in Figure 12.1.

Surface aerators, such as brush rotors, disc aerators, draft tube aerators, or fine bubble diffusers are used to circulate the mixed liquor. The mixing process entrains oxygen into the mixed liquor to foster microbial growth and the motive velocity ensures contact of microorganisms with the incoming wastewater. The aeration sharply increases the dissolved oxygen (DO) concentration but decreases as biomass uptake oxygen as the mixed liquor travels through the ditch. Solids are maintained in suspension as the mixed liquor circulates around the ditch. If design SRTs are selected for nitrification, a high degree of nitrification will occur. Oxidation ditch effluent is usually settled in a separate secondary clarifier. An anaerobic tank may be added before the ditch to enhance biological phosphorus removal.

An oxidation ditch may also be operated to achieve denitrification. One of the common design modifications for enhanced nitrogen removal is known as the Modified Ludzack-Ettinger (MLE) process (2, 4–8). In this process, illustrated in Figure 12.2, an anoxic tank is added upstream of the ditch along with mixed liquor recirculation from the aerobic zone to the tank to achieve higher levels of denitrification. In the aerobic basin, autotrophic bacteria (nitrifiers) convert ammonia-nitrogen to nitrite-nitrogen and then to nitrate-nitrogen. In the anoxic zone, heterotrophic bacteria convert nitrate-nitrogen to nitrogen gas which is released to the atmosphere. Some mixed liquor from the aerobic basin is recirculated to the anoxic zone to provide the mixed liquor with a high-concentration of nitrate-nitrogen to the anoxic zone.

Several manufacturers have developed modifications to the oxidation ditch design to remove nutrients in conditions cycled or phased between the anoxic and aerobic states. Although the mechanics of operation differ by manufacturer, in general, the process consists of two separate aeration basins, the first anoxic and the second aerobic. Wastewater and return activated sludge (RAS) are introduced into the first reactor which operates under anoxic conditions. Mixed liquor then flows into the second reactor operating under aerobic

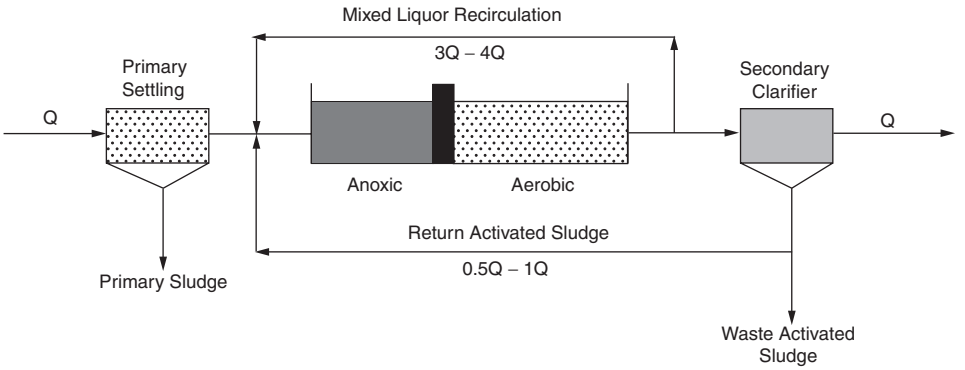

Fig. 12.2. The modified Ludzack-Ettinger process (2).

conditions. The process is then reversed and the second reactor begins to operate under anoxic conditions (2).

Another proposed configuration (9) is to obtain nitrification in the region just downstream of the brush aerators which is aerobic. As the liquor travels downstream and the oxygen is consumed, an anaerobic zone is formed. By routing a small portion of the raw sewage influent (as a carbon source) to this zone, denitrification occurs. The mixed liquor then contacts another brush aerator so that the organic nitrogen produced by the denitrifying bacteria is oxidized. The number of anaerobic zones and aerators required is a design parameter that depends on the capacity and loading of the plant.

3. APPLICABILITY

The oxidation ditch process is a fully demonstrated secondary wastewater treatment technology, applicable in any situation where activated sludge treatment (conventional or extended aeration) is appropriate (10). Oxidation ditches are applicable in plants that require nitrification because the basins can be sized using an appropriate SRT to achieve nitrification at the mixed liquor minimum temperature (11). This technology is very effective in small installations (wastewater flows between 0.1 and 10 MGD), small communities, and isolated institutions, because it requires more land than conventional treatment plants (2, 3).

The oxidation process as mentioned previously, originated in the Netherlands, with the first full scale plant installed in Voorschoten, Holland, in 1954. By the end of the century more than 9200 municipal oxidation ditch installations were operational in the United States (12). Nitrification to less than 1 mg/L ammonia-nitrogen consistently occurs when ditches are designed and operated for nitrogen removal. Today, a complete biological treatment system can be provided with a single oxidation ditch system. The oxidation ditch structure can be constructed with only a single aerator and an intrachannel clarifier. By incorporating denitrification within a channel of the oxidation ditch, alternating oxic/anoxic conditions can be created which will effectively reduce nitrogen concentrations to the desired low levels to meet the effluent discharge regulations (13). Double or triple concentric ditch arrangement allows for variation in dissolved oxygen levels resulting in conditions that are favorable for the biomass to remove nitrogen and phosphorus (14).

4. ADVANTAGES AND DISADVANTAGES

The main advantage of the oxidation ditch is the ability to achieve removal performance objectives with low operational requirements and operation and maintenance costs. Some specific advantages of oxidation ditches include (2):

(a) An added measure of reliability and performance over other biological processes owing to a constant water level and continuous discharge which lowers the weir overflow rate and eliminates the periodic effluent surge common to other biological processes, such as SBRs.
(b) Long hydraulic retention time and complete mixing minimize the impact of a shock load or hydraulic surge.

(c) Produces less sludge than other biological treatment processes owing to extended biological activity during the activated sludge process.
(d) Energy efficient operations result in reduced energy costs compared with other biological treatment processes.

The disadvantages include:

(a) Effluent suspended solids concentrations are relatively high compared to other modifications of the activated sludge process.
(b) Requires a larger land area than other activated sludge treatment options. This can prove costly, limiting the feasibility of oxidation ditches in urban, suburban, or other areas where land acquisition costs are relatively high.

5. DESIGN CRITERIA

Oxidation ditches are commonly constructed using reinforced concrete, although gunite, asphalt, butyl rubber, and clay have also been used. Impervious materials are usually used to prevent erosion. The ditches are usually 4 to 6 ft deep with 45 degrees or vertical sidewalls (3).

Screened wastewater enters the ditch, is aerated, and circulates at about 0.25 to 0.35 m/s (0.8 to 1.2 ft/s) to maintain the solids in suspension (15). The RAS recycle ratio is from 75 to 150%, and the mixed liquor suspended solids (MLSS) concentration ranges from 1500 to 5000 mg/L (15). The oxygen transfer efficiency of oxidation ditches ranges from 2.5 to 3.5 lb/hp-h (2, 16).

The design criteria are affected by the influent wastewater parameters and the required effluent characteristics, including the decision or requirement to achieve nitrification, denitrification, and/or biological phosphorus removal. Specific design parameters for oxidation ditches include (2).

5.1. Solids Retention Time (SRT)

Oxidation ditch volume is sized based on the required SRT to meet effluent quality requirements. The SRT is selected as a function of nitrification requirements and the minimum mixed liquor temperature. Design SRT values vary from 4 to 48 or more days (2, 3). Typical SRTs required for nitrification range from 12 to 24 days.

5.2. BOD Loading

BOD loading rates vary from less than 160 mg/L/d (10 lb/1000 ft^3/d) to more than 800 mg/L/d (50 lb/1000 ft^3/d) (2, 3). A BOD loading rate of 240 mg/L/d (15 lb/1000 ft^3/d) is commonly used as a design loading rate. However, the BOD loading rate is not typically used to determine whether or not nitrification occurs.

5.3. Hydraulic Retention Time

Although rarely used as a basis for oxidation ditch design, hydraulic retention times (HRTs) within the oxidation ditch range from 6 to 30 hours for most municipal wastewater treatment plants (2, 3).

6. PERFORMANCE

As fully demonstrated secondary treatment processes, oxidation ditch processes are readily adaptable for nitrification and denitrification. As part of an evaluation of oxidation ditches for nutrient removal (17), performance data were collected from 17 oxidation ditch plants. The average design flow for these plants varied between 378 and 45,425 m^3/d (0.1 to 12 MGD). The average performance of these plants indicates that oxidation ditches achieve BOD, suspended solids, and ammonia nitrogen removal of greater than 90%. Likewise, US EPA reported nitrogen removals of greater than 90% from several oxidation ditch processes (2).

It should be kept in mind that to be able to achieve such high nitrogen removals, it is imperative to have continuous plant supervision and skilled operation. This is essential for assuring full control of the dissolved oxygen (DO) profile in the oxidation ditch system. Several modeling techniques have been proposed to help for DO control and to perform real time predictions of performance (18, 19). The following sections discuss the performance of two recently designed oxidation ditch facilities.

6.1. Casa Grande Water Reclamation Facility

The City of Casa Grande, Arizona, Water Reclamation Facility began operation in February 1996. The system was designed to treat a wastewater flow of 15,142 m^3/d (4.0 MGD) and uses an anoxic zone preceding the aerobic zone of each train to provide denitrification. With influent design parameters of 270 mg/L BOD, 300 mg/L TSS, and 45 mg/L TKN, the plant has consistently achieved effluent objectives of 10 mg/L BOD, 15 mg/L TSS, 1.0 mg/L ammonia, and 5.0 mg/L nitrate-nitrogen. Table 12.1 summarizes the plant's performance between July 1997 and July 1999 (20).

6.2. Edgartown, Massachusetts WWTP

The Edgartown, Massachusetts WWTP, located on the island of Martha's Vineyard, is designed to treat 757 m^3/d (0.20 MGD) in the winter months and 2,839 m^3/d (0.75 MGD) in the summer. Two oxidation basins are installed and the plant has achieved performance objectives since opening. Table 12.2 summarizes average monthly influent, effluent and percent removal data (21).

Table 12.1
Performance of Casa Grande, AZ WWTP[a]

Parameter	BOD	TSS	Total N
Influent, average monthly value, mg/L	226	207	35
Effluent, average monthly value, mg/L	9	5	2
Removal, %	96	97	94

[a]Data adapted from ref. 20.

Table 12.2
Performance of Edgartown, MA WWTP[a]

Parameter	BOD	TSS	Total N
Influent, average monthly value, mg/L	238	202	27
Effluent, average monthly value, mg/L	3	5	2
Removal, %	99	97	92

[a]Data adapted from ref. 21.

7. PACKAGE OXIDATION DITCH PLANTS

Package plants are premanufactured treatment facilities used to treat wastewater in small communities. Package plants are usually designed by manufacturers to treat flows as low as 0.002 MGD to as high as 0.5 MGD (22, 23).

7.1. Description

Package oxidation ditches are typically manufactured in sizes that treat wastewater flow rates between 0.01 and 0.5 MGD. As seen in Figure 12.3, raw wastewater is first screened before entering the oxidation ditch. Depending on the system size and manufacturer type, a grit chamber may be required. Once inside the ditch, the wastewater is aerated with mechanical surface or submersible aerators (depending on manufacturer design) that propel the mixed liquor around the channel at velocities high enough to prevent solids deposition. The aerator ensures that there is sufficient oxygen in the fluid for the microbes and adequate mixing to ensure constant contact between the organisms and the food supply (24).

Treated sewage moves to the settling tank or final clarifier, where the biosolids and water separate. Wastewater then moves to other treatment processes while sludge is removed. Part of it is returned to the ditch as RAS, while the rest is removed from the process as the waste activated sludge (WAS). WAS is wasted either continuously or daily and must be stabilized before disposal or beneficial reuse.

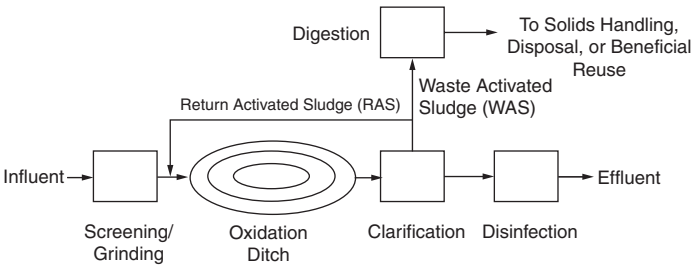

Fig. 12.3. Package oxidation ditch plant (22).

7.2. Applicability

In general, package treatment plants are applicable for areas with a limited number of people and small wastewater flows. They are most often used in remote locations such as trailer parks, highway rest areas, and rural areas. Oxidation ditches are suitable for facilities that require nutrient removal, have limitations owing to the nature of the site, or want a biological system that saves energy with limited use of chemicals unless required for further treatment.

Oxidation ditch technology can be used to treat any type of wastewater that is responsive to aerobic degradation. In addition, systems can be designed for denitrification and phosphorous removal. Types of industries using oxidation ditches include: food processing, meat and poultry packing, breweries, pharmaceutical, milk processing, petrochemical, and numerous other types. Oxidation ditches are particularly useful for schools, small industries, housing developments, and small communities. Ultimately, this technology is most applicable for places that have a large amount of land available (22).

7.3. Advantages and Disadvantages

Some advantages of package oxidation ditch plants are listed below (22):

(a) Systems are well-suited for treating typical domestic waste, have moderate energy requirements, and work effectively under most types of weather.
(b) Oxidation ditches provide an inexpensive wastewater treatment option with both low operation and maintenance costs and operational needs.
(c) Systems can be used with or without clarifiers, which affects flexibility and cost.
(d) Systems consistently provide high quality effluent in terms of TSS, BOD, and ammonia levels.
(e) Oxidation ditches have a relatively low sludge yield, require a moderate amount of operator skill, and are capable of handling shock and hydraulic loadings.

The disadvantages include:

(a) Oxidation ditches can be noisy owing to mixer/aeration equipment, and tend to produce odors when not operated correctly.
(b) Biological treatment is unable to treat highly toxic waste streams.
(c) Systems have a relatively large footprint.
(d) Systems have less flexibility should regulations for effluent requirements change.

7.4. Design Criteria

Key components of a typical oxidation ditch include a screening device, an influent distributor (with some systems), a basin or channel, aeration devices (mechanical aerators, jet mixers, or diffusers, depending on the manufacturer), a settling tank or final clarifier (with some systems), and an RAS system (with some systems). These components are often built to share a common wall to reduce costs and save space. Concrete tanks are typically used when installing package plant oxidation ditches. This results in lower maintenance costs as concrete tanks do not require periodic repainting or sand blasting. Fabricated steel or a combination of steel and concrete can also be used for construction, depending on site conditions (24).

Table 12.3
Design criteria for package oxidation ditch plants (22)

Parameter	Design value
BOD loading (F/M), lb BOD_5/lb MLVSS	0.05–0.30
Average oxygen requirement (@ 20°C), lb/lb BOD_5 applied	2–3
Peak Oxygen requirement (@ 20°C), lb/lb BOD_5 applied	1.5–2.0
MLSS, mg/L	3000–6000
Detention time, h	18–36
Volumetric loading, lb $BOD_5/1{,}000\,ft^3$	5–30

Table 12.4
Package oxidation ditch plants performance (22)

	Typical Effluent Quality		Ocoee WWTP	
	With 2° Clarifier	With Filter	% Removal	Effluent
CBOD, mg/L	0.10	5	>97	4.8
TSS, mg/L	0.10	5	>97	0.32
TP, mg/L	2	1	NA	NA
N-NO_3, mg/L	NA	NA	>95	0.25

2° = Secondary, NA = Not applicable.

Table 12.3 lists typical design parameters for package oxidation ditch plants. The volume of the oxidation ditch is determined based on influent wastewater characteristics, effluent discharge requirements, HRT, SRT, temperature, mixed liquor suspended solids (MLSS), and pH. It may be necessary to include other site specific parameters to design the oxidation ditch as well. Some oxidation ditches do not initially require clarifiers, but can later be upgraded and expanded by adding clarifiers, changing the type of process used, or adding additional ditches (25).

7.5. Performance

Although the manufacturer's design may vary, most oxidation ditches typically achieve the effluent limitations listed in Table 12.4. Denitrifying oxidation ditches are capable of extremely high efficiencies. With modifications, some oxidation ditches can achieve TN removal to 5 mg/L. The 3MGD oxidation ditch in Stonybrook, New York regularly maintains 97% nitrogen removal efficiency (9).

Currently, the wastewater treatment plant in Ocoee, Florida accepts an average flow of 1.1 to 1.2 MGD. The city chose to use an oxidation ditch because it was an easy technology for the plant staff to understand and implement. The facility is also designed for denitrification without the use of chemical additives. Nitrate levels consistently test at 0.8 to 1.0 mg/L with limits of 12 mg/L (26). Table 12.4 indicates how well the Ocoee oxidation ditch performs.

Table 12.5
Costs for package oxidation ditch plants* (22)

Flow range, MGD	Budget price, USD	Budget cost, USD/gal
0.00–0.03	96,000	6.39
0.03–0.06	109,100	2.42
0.06–1.10	116,300	0.21
1.10–1.70	126,500	0.10
1.70–2.50	138,100	0.07

*Dollars values adjusted form original 1999 (Cost Index = 460.16) to 2008 (Cost Index = 552.16); (Appendix A. extracted from US Army Corps of Engineers Ref. 27).

7.6. Costs

Table 12.5 lists budget cost estimates for various sizes of oxidation ditches (22). Operation and maintenance costs for oxidation ditches are significantly lower than other secondary treatment processes. In comparison to other treatment technologies, energy requirements are low, operator attention is minimal, and chemical addition is not required.

8. OPERATION AND MAINTENANCE

Oxidation ditches require relatively little maintenance compared to other secondary treatment processes. No chemicals are required in most applications, but metal salts can be added to enhance phosphorus removal.

8.1. Residuals Generated

Primary sludge is produced if primary clarifiers precede the oxidation ditch. Sludge production for the oxidation ditch process ranges from 0.2 to 0.85 kg TSS/kg (0.2 to 0.85 lb TSS/lb) BOD applied (28). Typical sludge production is 0.65 kg TSS/kg of BOD (0.65 lb TSS/lb of BOD). This is less than conventional activated sludge facilities because of long SRTs.

8.2. Operating Parameters

The oxygen coefficient for BOD removal varies with temperature and SRT. Typical oxygen requirements range from 1.1 to 1.5 kg of O_2 per kg of BOD removed (1.1 to 1.5 lb of O_2 per lb of BOD removed) and 4.57 kg O_2/kg TKN oxidized (4.57 lb O_2/lb TKN oxidized) (17). Oxygen transfer efficiency ranges from 2.5 to 3.5 lb/hp-h (16).

9. DESIGN CONSIDERATIONS

9.1. Input Data

The following data for flows and influent and effluent characteristics shall be provided (1):

(a) Wastewater flow (average and peak). In case of high variability, a statistical distribution should be provided.

(b) Wastewater strength
   1. BOD_5 (soluble and total), mg/L
   2. COD and/or *TOC* (maximum and minimum), mg/L
   3. Suspended solids, mg/L
   4. Volatile suspended solids, mg/L
   5. Nonbiodegradable fraction of VSS, mg/L

(c) Other characteristics
   1. pH
   2. Acidity and/or alkalinity, mg/L
   3. Nitrogen, mg/L (NH_3 or Kjeldahl)
   4. Phosphorus (total and soluble), mg/L
   5. Oils and greases, mg/L
   6. Heavy metals, mg/L
   7. Toxic or special characteristics (e.g., phenols), mg/L
   8. Temperature, °F or °C

(d) Effluent quality requirements
   1. BOD_5, mg/L
   2. *SS*, mg/L
   3. *TKN*, mg/L
   4. *P*, mg/L

9.2. Design Parameters

(a) Eckenfelder reaction rate constants and coefficients

$k = 0.0007$ to $0.002\,\text{L/mg/h}$

   1. $a = 0.73$
   2. $a' = 0.52$
   3. $b = 0.075/\text{d}$
   4. $b' = 0.15/\text{d}$
   5. $a_\text{o} = 0.77\text{a} = 0.56$
   6. $f = 0.140$
   7. $f' = 0.53$

(b) $F/M = 0.03 - 0.1$

(c) Volumetric loading = 10 to 40

(d) $t = 18$ to 36 h

(e) $t_\text{s} = 20$ to 30 d

(f) MLSS = 4000 to 8000 mg/L (mean = 6000 mg/L)

(g) *MLVSS* = 2800 to 5600 mg/L

(h) $Q_\text{r}/Q = 0.5$ to 1.0

(i) lb O_2/lb $BOD_\text{r} \geq 1.5$

(j) lb solids/lb $BOD_\text{r} \leq 0.2$.

(k) $\theta = 1.0$ to 1.03

(l) Efficiency $\geq 90\%$

9.3. Design Procedure

The following is a guide line that summarizes the design procedure (Eckenfelder Method) for an oxidation ditch (1, 29–35)

(a) Assume the following design parameters when known.
1. Fraction of BOD synthesized (a)
2. Fraction of BOD oxidized for energy (a')
3. Endogenous respiration rate (b and b')
4. Fraction of BOD_5 synthesized to degradable solids (a_o)
5. Nonbiodegradable fraction of *VSS* in influent (f)
6. Mixed liquor suspended solids (*MLSS*)
7. Mixed liquor volatile suspended solids (*MLVSS*)
8. Temperature correction coefficient (θ)
9. Degradable fraction of the *MLVSS* (x')
10. Food-to-microorganism ratio (F/M)
11. Effluent soluble BOD_5 (S_e)

(b) Adjust the BOD removal rate constant for temperature

$$k_T = k_{20}\theta^{(T-20)} \tag{1}$$

where
k_T = rate constant for desired temperature
k_{20} = rate constant at 20°C
θ = temperature correction coefficient
T = temperature, °C

(c) Determine the size of the aeration tank

$$V = a_o(S_o - S_e)Q_{avg}/X_V f'b \tag{2}$$

where
V = aeration tank volume, MG
a_o = fraction of BOD_5 synthesized to degradable solids
S_o = influent BOD_5, mg/L
S_e = effluent soluble BOD_5, mg/L
Q_{avg} = Average waste flow, MGD
X_V = MLVSS, mg/L
f' = degradable fraction of the MLVSS
b = endogenous respiration rate, 1/d

(d) Calculate the detention time

$$t = (V/Q)24 \tag{3}$$

where
t = detention time, h
V = volume, MG
Q = flow, MGD

(e) Assume the organic loading and calculate detention time

$$t = (24S_o)/X_V(F/M) \tag{4}$$

where
t = detention time, h
S_o = influent BOD_5, mg/L

X_V = volatile solids in raw sludge, mg/L
F/M = organic loading (food-to-microorganism ratio)
and select the larger of two detention times from d or e above

(f) Determine the oxygen requirements allowing 60% for nitrification during summer

$$O_2 = [a' S_r Q_{avg} + b' X_V V + 0.6(4.57)(TKN)(Q_{avg})] \quad (8.34) \tag{5}$$

where
O_2 = oxygen required, lb/d
a' = fraction of BOD oxidized for energy
S_r = BOD_5 removed, mg/L
Q_{avg} = average waste flow, MGD
b' = endogenous respiration rate, 1/d
X_V = MLVSS, mg/L
V = aeration tank volume, MG
4.57 = parts oxygen required per part TKN
TKN = total Kjeldahl nitrogen, mg/L

(g) calculate oxygen requirement per lb BOD_r (it should be ≥ 1.5)

$$\text{lb}O_2/\text{lb BOD}_r = O_2/Q_{avg}\ S_r \quad (8.34) \tag{6}$$

where
O_2 = oxygen required, lb/d
Q_{avg} = average wastewater flow, MGD
S_r = BOD_5 removed, mg/L

(h) Calculate sludge production

$$\Delta X_V = 8.34[a(S_r)(Q) - (b)(X_V)(V) - Q(SS)_{eff} + Q(VSS)f' + Q(SS - VSS)] \tag{7}$$

where
ΔX_V volatile sludge produced, lb/d
a = fraction of BOD synthesized
S_r = BOD_5 removed, mg/L
Q = average wastewater flow, MGD
b = endogenous respiration rate, 1/d
X_V = volatile solids in raw sludge, mg/L
V = aeration tank volume, MG
$(SS)_{eff}$ = effluent suspended solids, mg/L
VSS = volatile suspended solids in influent, mg/L
f' = degradable fraction of the MLVSS
SS = suspended solids in influent, mg/L

(i) Calculate solids produced per pound of BOD removed (it should be ≥ 1.5)

$$\text{lb solids/lb BOD}_r = \Delta X_V/Q(S_o - S_e)8.34 \tag{8}$$

where
ΔX_V = volatile sludge produced, lb/d
Q = waste flow, MGD

S_o = influent BOD_5, mg/L
S_e = effluent soluble BOD_5, mg/L

(j) Calculate the solids retention time

$$t_s = X_a V(8.34)/\Delta X_V \tag{9}$$

where
t_s = solids retention time, d
X_a = MLSS, mg/L
V = volume of aeration tank, MG
ΔX_V = volatile sludge produced, lb/d

(k) Determine the effluent soluble BOD_5

$$S_e/S_o = 1/1 + kX_V t \tag{10}$$

where
S_e = soluble effluent BOD, mg/L
S_o = influent BOD_5, mg/L
k = rate constant, L/mg/h
X_V = MLVSS, mg/L
t = aeration time, h

(l) Calculate sludge recycle ratio

$$Q_r/Q_{avg} = X_a/X_u - X_a \tag{11}$$

where
Q_r = volume of recycled sludge, MGD
Q_{avg} = average flow, MGD
X_a = MLSS, mg/L
X_u = suspended solids concentration in returned sludge, mg/L

(m) Calculate the nutrient requirements for nitrogen and Phosphorus

$$N = 0.123\Delta X_V \tag{12}$$

$$P = 0.026\Delta X_V \tag{13}$$

where
ΔX_V = sludge produced, lb/d

9.4. Output Data

(a) Aeration Tank (1)
1. Reaction rate constant, L/mg/h
2. Sludge produced per BOD removed
3. Endogenous respiration rate (b, b')
4. O_2 used per BOD removed
5. Influent nonbiodegradable VSS
6. Effluent degradable VSS

7. lb BOD/lb MLSS-d (F/M)
8. Mixed liquor suspended solids (MLSS), mg/L
9. Mixed liquor volatile suspended solids (MLVSS), mg/L
10. Aeration time, h
11. Volume of aeration tank, MG
12. Oxygen required, lb/d
13. Sludge produced, lb/d
14. Nitrogen requirement, lb/d
15. Phosphorus requirement, lb/d
16. Sludge recycle ratio
17. Solids retention time, d

(b) Mechanical Aeration System
1. Standard transfer efficiency, lb O_2/hp-h
2. Operating transfer efficiency, lb O_2/hp-h
3. Horsepower required, hp

(c) Diffused Aeration System
1. Standard transfer efficiency, %
2. Operating efficiency, %
3. Required air flow, cfm/1000 ft^3

10. COSTS

The basin volume and footprint required for oxidation ditch plants have traditionally been very large compared with other secondary treatment processes. Larger footprints result in higher capital costs, especially in urbanized locations where available land is very expensive. Vertical reactors, in which process flow travels downward through the reactor, are generally more expensive than traditional horizontal reactors. However, because they require less land than more conventional horizontal reactors, they can significantly reduce overall capital costs where land costs are high.

The cost of an oxidation ditch plant varies depending on treatment capacity size, design effluent limitations, land cost, local construction costs, and other site specific factors. Construction capital costs for ten plants were evaluated by US EPA in 1991 (17), with construction costs, in 2008 Dollars, ranging from USD 0.73 to 4.46/L/d (USD 2.76 to 16.87/gpd) treated. The cost values have been adjusted from the original 1991 (Cost Index 392.35) to 2008 (Cost Index 552.16) using the Utilities Cost index (Appendix A. Ref. 27).

Recent information obtained from manufacturers on facilities ranging 3,785 to 25,740 m^3/d (1.0 MGD to 6.8 MGD) indicates that construction capital costs (adjusted from original 1999 to 2008 Dollars) of oxidation ditch plants range from USD 0.80 to 1.32/L/d (USD 3.00 to 4.80/gpd). For example, the Blue Heron Water Reclamation Facility in Titusville, Florida (36) a 15,142 m^3/d (4.0 MGD) oxidation ditch and sludge handling facility which began operation in 1996, was constructed for about USD 0.96/L/d (USD 3.60/gpd). The facility features a multi-stage biological nutrient removal process and a sophisticated Supervisory Control and Data Acquisition System (SCADA) control system.

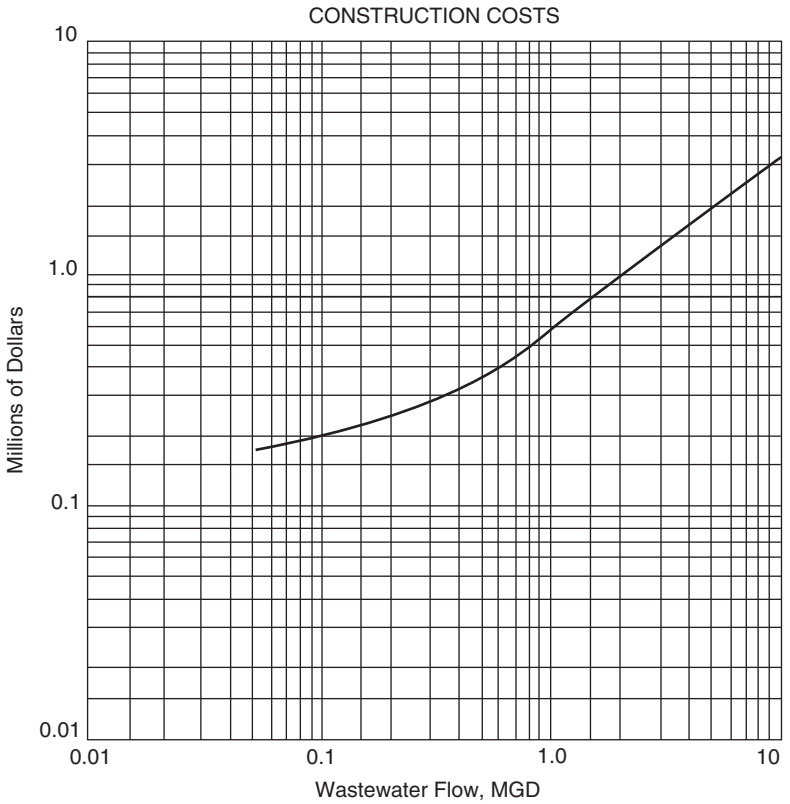

Fig. 12.4. Oxidation ditch construction costs (3).

Construction costs (1976 Dollars, Utilities Index = 202.61) for oxidation ditches are shown in Figure 12.4. To obtain the values in terms of the 2008 US Dollars, using the Cost Index for Utilities (Appendix A), multiply the costs by a factor of 552.16/202.61 = 2.72 (27). Construction costs include oxidation ditch, clarifier, pumps, building, out-door sludge drying beds, laboratory, but excludes land, engineering, legal and financing during construction.

Oxidation ditches offer significantly lower operation and maintenance (O&M) costs than other secondary treatment processes. Compared to other treatment technologies, energy requirements are low, operator attention is minimal, and chemical addition is not usually required. For example the Tar River Wastewater Reclamation Facility in Louisburg, North Carolina has documented energy savings of 40% compared with conventional activated sludge plants (37). The oxidation ditch has also eliminated chemical costs and plant staff is available for other facility needs (37).

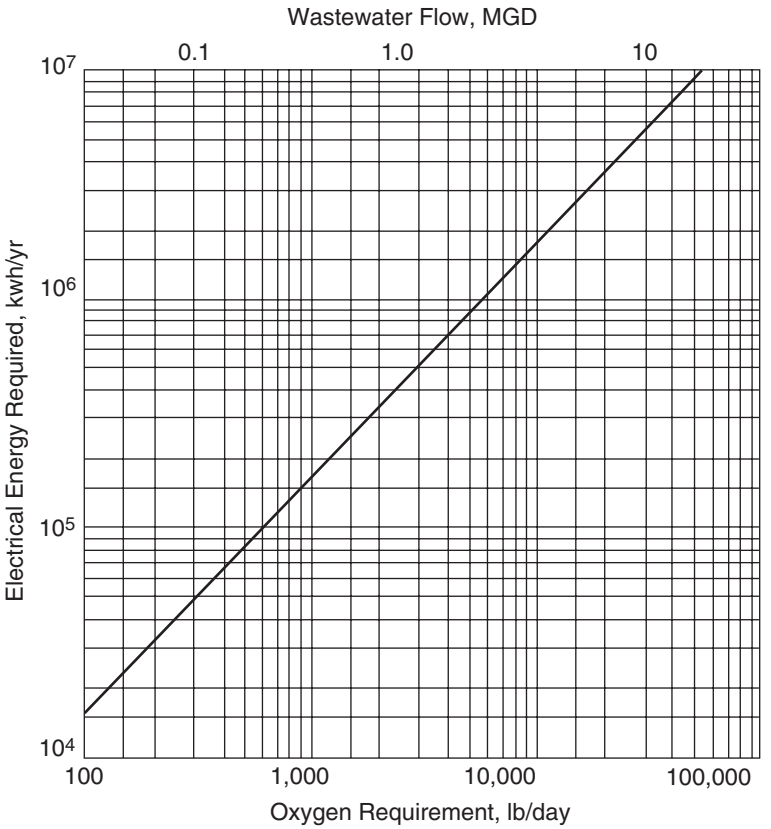

Fig. 12.5. Oxidation ditch energy requirements (3).

Energy requirements and O&M costs are shown in Figures 12.5 and 12.6 respectively. O&M costs are given in 1976 Dollars, Utilities Index = 202.61. To obtain the values in terms of 2008 US Dollars, using the Cost Index for Utilities (Appendix A), multiply the costs by a factor of 552.16/202.61 = 2.72 (27).

Energy requirements are based on the following assumptions:

(a) Type of Energy Required: Electrical
(b) Influent BOD_5 = 136; Effluent BOD_5 = 20 mg/L
(c) Oxygen transfer efficiency = 1.8 lb O_2/hp-h (wire to water)
(d) No appreciable nitrification occurs
(e) Oxygen requirement = 1.5 lb O_2/lb BOD_5 removed

O&M costs include labor, utilities, chemicals, maintenance and materials. For further details, the reader is referred to the extensive literature dealing with this subject (12, 15, 30, 38–40).

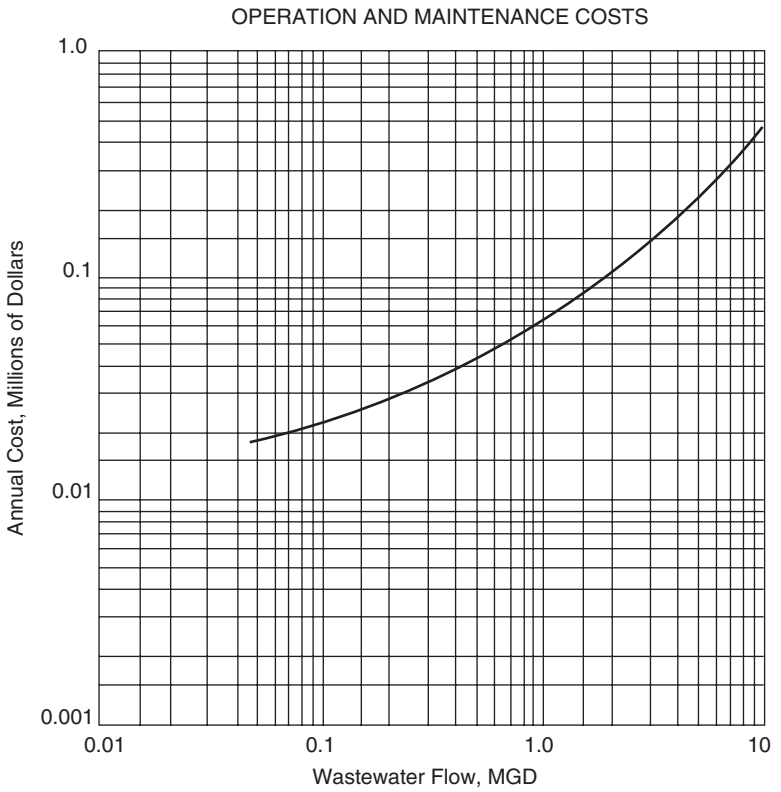

Fig. 12.6. Oxidation ditch operation and maintenance costs (3).

11. DESIGN EXAMPLE

Design an oxidation ditch for the treatment of the following wastewater (1):

Average wastewater flow = 1.0 MGD
Wastewater $BOD_5 = 200\,mg/L$
Wastewater $SS = 200\,mg/L$
Wastewater $VSS = 150\,mg/L$
Total organic nitrogen concentration = 30 mg/L
Total phosphorus concentration = 15 mg/L
Required effluent $BOD_5 = 10\,mg/L$
Required effluent $SS = 20\,mg/L$
Required underflow sludge concentration = 10,000 mg/L
Wastewater temperature = 15°C

(a) Given or assume the following design parameters:
1. Fraction of BOD synthesized, $a = 0.73$
2. Fraction of BOD oxidized for energy, $a' = 0.52$
3. Endogenous respiration rate, $b = 0.075/d$; and $b' = 0.15/d$

4. Fraction of BOD_5 synthesized to degradable solids, $a_o = 0.56$
5. Nonbiodegradable fraction of VSS in influent, $f = 0.40$
6. Mixed liquor suspended solids, $MLSS = X_a = 6000\,mg/L$
7. Mixed liquor volatile suspended solids, $MLVSS = X_V = 4200\,mg/L$
8. Temperature correction coefficient, $\theta = 1.02$
9. Degradable fraction of the *MLVSS*, $f' = 0.53$
10. Food-to-microorganism ratio, $F/M = 0.06$
11. Effluent soluble BOD_5, $S_e = 10\,mg/L$

(b) Adjust the BOD removal rate constant for temperature

$$k_T = k_{20}\theta^{(T-20)}$$

where
k_T = rate constant for desired temperature
k_{20} = rate constant at 20°C = 0.0010
θ = temperature correction coefficient = 1.02
T = temperature, 15°C

$$k_T = 0.0010(1.02)^{15-20}$$

$$k_T = 0.0009$$

(c) Determine the size of the aeration tank

$$V = a_o(S_o - S_e)Q_{avg}/X_V f' b$$

where
V = aeration tank volume, MG
a_o = fraction of BOD_5 synthesized to degradable solids = 0.56
S_o = influent BOD_5, 200 mg/L
S_e = effluent soluble BOD_5, 10 mg/L
Q_{avg} = Average waste flow, 1.0 MGD
X_V = MLVSS, 4200 mg/L
f' = degradable fraction of the MLVSS = 0.53
b = endogenous respiration rate, 0.075/d

$$V = 0.56(200 - 10)1.0/4200(0.53)0.075$$

$$V = 0.64\,MG$$

(d) Calculate the detention time

$$t = (V/Q)24$$

where
t = detention time, hr
V = volume, 0.64 MG
Q = average flow, 1.0 MGD

$$t = (0.64/1.0)24$$

$$t = 15\,h$$

(e) Assume the organic loading and calculate detention time

$$t = (24S_o)/X_V(F/M)$$

where
t = detention time, hr
S_o = influent BOD_5, 200 mg/L
X_V = volatile solids in raw sludge, 4,200 mg/L
F/M = organic loading (food-to-microorganism ratio) = 0.06

$$t = 24(200)/4{,}200(0.06)$$

$$t = 19\,\text{h}$$

Select the larger of two detention times from d or e above; $t = 19$ h

(f) Determine the oxygen requirements allowing 60% for nitrification during summer

$$O_2 = a'S_r\,Q(8.34) + b'\,X_V V(8.34) + 0.6(4.57)(TKN)(Q)(8.34)$$

where
O_2 = oxygen required, lb/d
a' = fraction of BOD oxidized for energy = 0.56
S_r = BOD_5 removed = $S_o - S_e = 200 - 10 = 190$ mg/L
Q = average waste flow, 1.0 MGD
b' = endogenous respiration rate, 0.15/d
X_V = MLVSS, 4200 mg/L
V = aeration tank volume, 0.64 MG
TKN = total Kjeldahl nitrogen, 30 mg/L

$$O_2 = [0.56(190)1.0 + 0.15(4{,}200)0.64 + 0.6(4.57)30(1.0)]8.34$$

$$O_2 = 4{,}936\,\text{lb/d}$$

(g) calculate oxygen requirement per lb BOD_r (it should be ≥ 1.5)

$$\text{lb}O_2/\text{lb BOD}_r = O_2/\text{Q S}_r(8.34)$$

where
O_2 = oxygen required, 4936 lb/d
Q = average wastewater flow, 1.0 MGD
S_r = BOD_5 removed, 190 mg/L

$$\text{lb O}_2/\text{lb BOD}_r = 4936/190(1.0)8.34$$

$$= 3.0 \geq 1.5\ \text{OK}$$

(h) Calculate sludge production

$$\Delta\,X_V = 8.34[\text{a}(S_r)(Q) - (\text{b})(X_V)(V) - Q(SS)_{\text{eff}} + Q(VSS)\,f' + Q(SS - VSS)]$$

where
$\Delta\,X_V$ volatile sludge produced, lb/d
a = fraction of BOD synthesized = 0.73

S_r = BOD_5 removed, 190 mg/L
Q = average wastewater flow, 1.0 MGD
b = endogenous respiration rate, 0.75/d
X_V = volatile solids in raw sludge, 4,200 mg/L
V = aeration tank volume, 0.64 MG
$(SS)_{eff}$ = effluent suspended solids, 20 mg/L
VSS = volatile suspended solids in influent, 150 mg/L
f' = degradable fraction of the MLVSS = 0.53
SS = suspended solids in influent, 200 mg/L

$$\Delta X_V = 8.34[0.73(190)1.0 - 0.075(4{,}200)0.64 - 1.0(20) + 1.0(150)0.53$$
$$+1.0(200 - 150)]8.34$$
$$\Delta X_V = 389 \text{ lb/d}$$

(i) Calculate solids produced per pound of BOD removed

$$\text{lb solids/lb } BOD_r = \Delta X_V / Q(S_o - S_e)8.34$$

where
ΔX_V = volatile sludge produced, 389 lb/d
Q = average flow, 1.0 MGD
S_o = influent BOD_5, 200 mg/L
S_e = effluent soluble BOD_5, 10 mg/L

$$\text{lb solids/lb } BOD_r = 389/1.0(200 - 10)8.34$$
$$\text{lb solids/lb } BOD_r = 0.25 \approx 0.2 \text{ OK}$$

(j) Calculate the solids retention time

$$t_s = X_a V(8.34)/\Delta X_V$$

where
t_s = solids retention time, d
X_a = MLSS, 6,000 mg/L
V = volume of aeration tank, 0.64 MG
ΔX_V = volatile sludge produced, 389 lb/d

$$t_s = 6000(0.64)8.34/389$$
$$t_s = 82 \text{ days}$$

(k) Determine the effluent soluble BOD_5

$$S_e/S_o = 1/1 + kX_V t$$

where
S_e = soluble effluent BOD, mg/L
S_o = influent BOD_5, 200 mg/L
k = rate constant, 0.0009 L/mg/h

X_V = MLVSS, 4,200 mg/L
t = aeration time, 19 h

$$S_e/200 = 1/1 + (0.0009)4{,}200(19)$$

$$S_e = 2.7\ \text{mg/L}$$

(l) Calculate sludge recycle ratio

$$Q_r/Q = X_a/X_u - X_a$$

where
Q_r = volume of recycled sludge, MGD
Q = average flow, 1.0 MGD
X_a = MLSS, 6000 mg/L
X_u = suspended solids concentration in returned sludge, mg/L = underflow concentration, 10,000 mg/L

$$Q_r/Q = 6000/10{,}000 - 6000$$

$$\text{Recycle ratio } Q_r/Q = 1.5$$

$$\text{and recycle flow, } Q_r = 1.5Q = 1.5\ \text{MGD}$$

(m) Calculate the nutrient requirements for nitrogen and Phosphorus

$$N = 0.123\Delta X_V$$

$$P = 0.026\Delta X_V$$

where
ΔX_V = sludge produced, 389 lb/d

$$N = 0.123(389) = 48\ \text{lb/d}$$

$$N \text{ in influent } = 30\ \text{mg/L}(1.0)8.34 = 250\ \text{lb/d} > 48$$

OK no N need to be added

$$P = 0.026(389)$$

$$P = 10\ \text{lb/d}$$

$$P \text{ in influent} = 15\ \text{mg/L}(1.0)8.34 = 125\ \text{lb/d} > 10\ \text{lb/d}$$

OK no P need to be added

NOMENCLATURE

4.57 = parts oxygen required per part TKN
a = fraction of BOD synthesized
a' = fraction of BOD oxidized for energy
a_o = fraction of BOD_5 synthesized to degradable solids
b = endogenous respiration rate, 1/d
b' = endogenous respiration rate, 1/d

f' = degradable fraction of the MLVSS
F/M = organic loading (food-to-microorganism ratio)
k = rate constant, L/mg/hr
k_{20} = rate constant at 20°C
k_T = rate constant for desired temperature
O_2 = oxygen required, lb/d
Q = waste flow, MGD
Q_{avg} = average wastewater flow, MGD
Q_r = volume of recycled sludge, MGD
S_e = effluent soluble BOD_5, mg/L
S_o = influent BOD_5, mg/L
S_r = BOD_5 removed, mg/L
SS = suspended solids in influent, mg/L
$(SS)_{eff}$ = effluent suspended solids, mg/L
T = temperature, °C
t = aeration time, hr
t_s = solids retention time, d
V = aeration tank volume, MG
VSS = volatile suspended solids in influent, mg/L
X_a = MLSS, mg/L
X_u = suspended solids concentration in returned sludge, mg/L
X_V = volatile solids in raw sludge, or MLVSS, mg/L
ΔX_V = volatile sludge produced, lb/d
θ = temperature correction coefficient

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APPENDIX

United States Yearly Average Cost Index for Utilities U.S. Army Corps of Engineers (27)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

13
Biological Nitrification and Denitrification Processes

Yue-Mei Lin, Joo-Hwa Tay, Yu Liu, and Yung-Tse Hung

CONTENTS

Abstract If the wastewater to be treated contains various forms of nitrogen, three biological treatment steps are required for nitrogen removal: (a) in a bio-oxidation step, organic nitrogen is anerobically broken down to ammonia nitrogen; (b) in a subsequent nitrification step, ammonia nitrogen in the wastewater is aerobically converted to nitrate nitrogen; and (c) in a final denitrification step, nitrate nitrogen is anaerobically or anoxically converted to nitrogen gas. This chapter discusses bio-oxidation, nitrification and denitrification process steps, their principles, and design considerations in detail.

Key Words Biological process•nitrogen removal•bio-oxidation•nitrification•denitrification•theory•design•process control•design.

1. INTRODUCTION

It is well known that aquatic organisms such as fish are sensitive to the presence of ammonia, whereas increased nitrogen loads in the effluent can stimulate the growth of aquatic plants and promote eutrophication. On the other hand, a high nitrate concentration in surface

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or ground water will cause infant methaemoglobinaemia if these are used as water supplies. Therefore, the removal of ammonium-nitrogen from water and wastewater has attained greater significance in recent years as a means of protecting and preserving the environment. To meet the increasingly stringent environmental regulations for nitrogen discharge, many treatment processes have been employed, such as biological processes, physico-chemical units, breakpoint-chlorination, ion-exchange, membrane processes, and precipitation. Within these treatment techniques, biological processes provide the most economical means for controlling nitrogen in wastewater effluents (1–9). The widely used biological nitrification and denitrification processes for nitrogen removal can be classified into two main categories: suspended and attached culture systems. Owing to the sensitivity of nitrifying bacteria to environmental factors as well as their lower growth rates, it is difficult to obtain and maintain a sufficient amount of nitrifying bacteria in conventional wastewater treatment plants. To address these problems, several modified and innovative nitrification and denitrification processes have been successfully developed in the past few years. This chapter will provide a review of the fundamentals of nitrification and denitrification as well as the latest technical developments in biological nitrogen removal processes.

2. FUNDAMENTALS OF NITRIFICATION

Nitrification is the biological formation of nitrate from sequentially oxidizing ammonium with the intermediate formation of nitrite. These oxidative reactions are catalyzed by two mutually exclusive groups of microorganisms, namely ammonium oxidizers and nitrite oxidizers.

2.1. Stoichiometry

The stoichiometric equation for the oxidation of ammonium to nitrite by ammonium oxidizer is as follows:

$$NH_4^+ + 1.5\,O_2 \rightarrow NO_2^- + 2\,H^+ + H_2O \tag{1}$$

This stoichimetric equation shows that the oxidation of ammonium to nitrite yields a release of free energy between 58 and 84 kcal per mole of ammonium (2). The reaction for the oxidation of nitrite to nitrate by nitrite oxidizer is:

$$NO_2^- + 1.5\,O_2 \rightarrow NO_3^- \tag{2}$$

The free energy released from this reaction has been estimated as between 15.4 and 20.9 kcal per mole of nitrite (3). Ammonium oxidizers thus obtain more energy than nitrite oxidizers from the oxidation of per mole of nitrogen. Using the empirical formula $C_5H_7NO_2$ for the formation of biomass, the following reactions have been proposed for the synthesis of the ammonium oxidizer and nitrite oxidizer, respectively:

$$55\,NH_4^+ + 76\,O_2 + 109\,HCO_3^- \rightarrow C_5H_7NO_2 + 54\,NO_2^- + 57\,H_2O + 104\,H_2CO_3 \tag{3}$$

$$400\,NO_2^- + NH_4^+ + 4\,H_2CO_3 + HCO_3^- + 195\,O_2 \rightarrow C_5H_7NO_2 + 3\,H_2O + 400\,NO_3^- \tag{4}$$

Combining the above stoichiometric equations for energy releasing and cell synthesis, the observed growth yield of nitrifying bacteria is in terms of the mass of volatile suspended solids (VSS) produced/mass of ammonium or nitrite oxidized. The theoretical yield coefficients are 0.29 g VSS/g NH_4^+-N and 0.084 g VSS/g NO_2-N, but values obtained from experiments are lower because a fraction of free energy released by oxidation is diverted to microbial maintenance functions (3). Based on the growth yields of 0.08 g VSS/g NH_4^+-N for ammonium oxidizer and 0.05 g VSS/g of NO_2-N for nitrite oxidizer, the overall reaction for complete nitrification can be written as (4):

$$NH_4^+ + 1.83\,O_2 + 1.98\,HCO_3^- \longrightarrow 0.021\,C_5H_7NO_2 + 1.041\,H_2O + 0.98\,NO_3^- + 1.88\,H_2CO_3 \quad (5)$$

This equation shows that (a) nitrification is an obligatorily aerobic process, and 4.18 g oxygen is required for the oxidation of one gram ammonium-nitrogen; (b) the overall growth yield of nitrifying bacteria can be calculated as 0.13 g biomass per gram ammonium-nitrogen oxidized, which is much lower than that of heterotrophic bacteria; (c) the oxidation of ammonium-nitrogen produces hydrogen ions in a ratio of 1 mole to 1 mole, as a result, approximately 8.62 g HCO_3^- per gram of NH_4^+-N oxidized is required to buffer the system against the hydrogen ions released from nitrification process.

2.2. Metabolism

All microorganisms can be placed in one of several nutritional categories based on their requirements for carbon, energy, and hydrogen atoms or electrons. Microorganisms typically associated with nitrification belong to a group called chemolithotrophics (chemolithotrophic autotrophs) that oxidize reduced inorganic compounds, such as iron, nitrogen or sulfur molecules to derive both energy and electrons for biosynthesis, and fix carbon dioxide as their carbon source via a special metabolic pathway called the Calvin cycle. The free energy released from the oxidation of both ammonia and nitrite is used to form adenosine 5′-triphosphate (ATP). This energy will be mainly used for the synthesis of microorganisms and the fixation of CO_2 including NADH or NADPH formation (Figure 13.1). The incorporation of one mole of carbon dioxide into organic material (glucose) requires three moles of ATP and two moles of NADPHs:

$$6CO_2 + 18ATP + 12NADPH + 12H^+ + 12H_2O \rightarrow \text{glucose} + 18ADP + 18P_i + 12NADP^+ \quad (6)$$

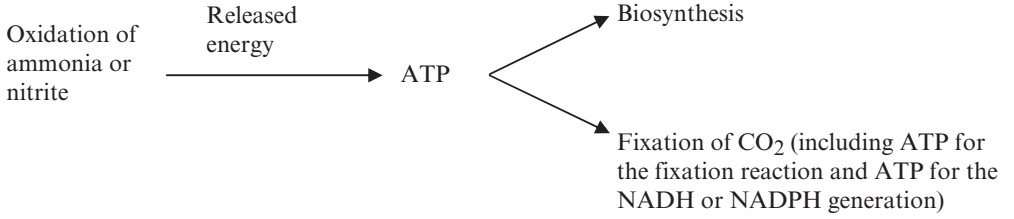


Fig. 13.1. Utilization of energy released by the oxidation of ammonia or nitrite.

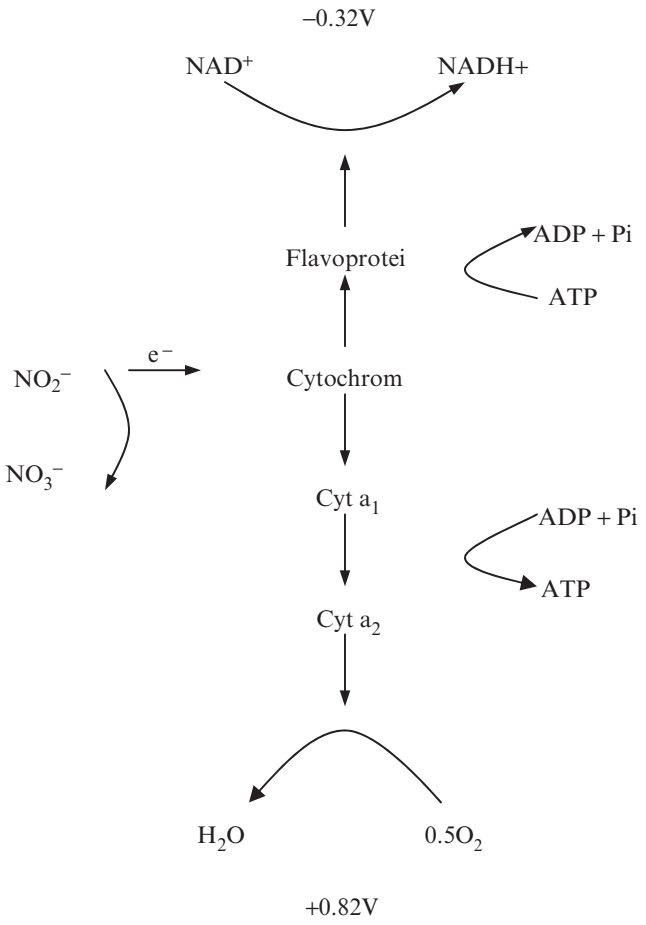


Fig. 13.2. Electron Flow of *Nitrobacter*. In the flow of electrons in the transport chain of *Nitrobacter*, electrons flow from nitrite to oxygen (down the reduction potential gradient). This process releases energy. But for the flow in the reverse direction from nitrite to NAD^+, protonmotive force or ATP energy is required to force electrons up the reduction potential gradient (adapted from (10)).

For this oxidation-reduction reaction involving electron transportation in cells, the participation of electron carriers such as NADH or NADPH is required. Because molecules such as ammonia and nitrite have more positive reduction potentials than NAD^+, they cannot directly donate their electrons to form the required NADH and NADPH as electrons spontaneously move only from donors with more negative reduction potentials to acceptors with more positive potentials. However, nitrifying bacteria solves this problem by using ATP to reverse the flow of electrons from nitrogen donors (Figure 13.2) (10). Therefore, considerable energy is used to generate NADH or NADPHs as well as ATP for the fixation of dioxide. For example, up to 80% of the energy produced by nitrite oxidation is used in the reduction of carbon dioxide.

Compared to the metabolism of heterotrophs, much less energy is available from the oxidation of inorganic molecules than from the complete oxidation of organic molecules.

This situation makes the net amount of energy for nitrifiers to grow and reproduce become much less than results in lower growth efficiencies of nitrifying bacteria. This situation can, on the other hand, push nitrifiers to oxidize a large quantity of inorganic material to meet their growing and reproducing requirements, in turn magnifying their ecological impact in a significant way.

2.3. Methods for Nitrifier Identification

In recent years, a variety of analytical methods using molecular microbiology techniques have been developed to analyze composition, architecture and physico-chemical properties of microbial community (11). As to nitrifying bacteria, they are closely related by their specialized biochemical reactions that oxidize reduced nitrogenous compounds for energy and fixing carbon dioxide for their carbon source. They can be classified by their cell shape, membrane constituents and G + C content. Note that the ammonium-oxidizers are distinguished by the prefix *Nitroso-* for the genus name, whereas the nitrite-oxidizers have the comparable prefix *Nitro-*.

To date, a total of 25 species of ammonia-oxidizers and eight species of nitrite-oxidizers are cultured, and the existence of many more species has been indicated by molecular in-situ investigations. It should be pointed out that only a portion of the existing nitrifiers has been defined via isolation and subsequent physiological/molecular characterization. Furthermore, the distribution patterns of the distinct species of nitrifying bacteria are closely dependent upon various environmental parameters. Hence, the composition of nitrifying bacterial communities is complex and diverse in heterogeneous habitats. It seems that isolation and characterization of as many as possible new species seems to be one of the most important points to be advanced. Nitrifying bacteria have slow growth rates, and are very sensitive to toxic shocks, pH and temperature swings. Difficulties associated with culture of these slow-growth species involved in nitrification have resulted in the development of indirect culture methods and procedures for direct observation of cells within the solid sample or in extracts. Traditionally, the two most commonly employed techniques for quantifying nitrifiers are the most-probable-number method (MPN) using nitrifier-specific growth media, and direct enumeration of specific organisms using fluorescent antibodies.

When the MPN method is used in isolation of nitrifying bacteria, one must be aware of its technical limitations. At first, as culture media used are selective, only a portion of the nitrifying population is able to grow. Secondly, the normal incubation time in nitrifier assays is much shorter than what is actually required for its growth. This would not allow them to reach maximal population development (12). But longer incubation periods may result in delays in data acquisition and the evaporation of the liquid growth media that may itself induce artifacts in population selection. Finally, it should be pointed out that the low statistical precision would enhance the limitations of MPN method that precludes use of the procedure for studies where detection of slight changes in population densities is necessary.

The fluorescent antibody technique was developed to overcome the problems inherent in the MPN procedure. The advantages of this method can be seen such that data are produced rapidly, specific populations are quantified, and small changes in population of the nitrifiers are easily measured. However, some technical problems are encountered in real application of the fluorescent antibody technique. Because many serotypes of autotrophic nitrifiers are

present, large numbers of fluorescent antibody types may have to be used to quantify the total nitrifier population; and it is difficult to isolate and purify nitrifiers, a prerequisite for antibody production. As a result, noncultured nitrifiers will not be detected by the fluorescent antibody method. Therefore, to better understand the mechanisms and establish stable nitrification, specific and rapid identification methods for nitrifiers are strongly required.

The development and application of recombinant DNA technique in domestic and industrial wastewater treatment is revolutionizing the ability to elucidate the structure and diversity of complex microbial ecosystems. This technique is based on the principles of specific nucleic acid probes and polymerase chain reaction. It allows identification of microbial types without the bias or tedium of culturing by targeting the cells' genetic code. Nucleic acid probe is a gene probe which is made of a piece of DNA controlling a desirable function in a cell and labeled with a radioactive element such as ^{32}P or with an enzyme (e.g., β-galactosidase, alkaline phosphatase). The probe can hybridize with a complementary strand of target DNA isolated from a given environmental sample or a bacterial colony (13, 14). Fluorescently labeled ribosomal RNA probes (rRNA probes) have been commonly used for the identification and classification of indigenous microorganisms in environmental samples (11, 15). Their sensitivity is much greater than that of DNA probes; however, indigenous bacteria often have a lower number of copies of ribosomes than cultured bacteria. Thus signal (e.g., fluorescence) amplification is sometimes needed.

Polymerase chain reaction, namely the PCR technique, was initially developed by Mullis et al. (16). In fact, this technique is essentially based on the DNA duplication process in vitro, and millions of copies of the target DNA sequence are created. Usually 30 cycles of DNA replication is required to make the target DNA fragment amplified and accumulated exponentially, and this only takes about 3 hours, which is much shorter than the time used for traditional culture tests. So far, the PCR technique has been automated by using a DNA thermal cycler to control the temperature necessary for the denaturation and annealing steps. The use of PCR technique greatly enhances the sensitivity of nucleic acid probes. In fact, some of these probes combined with PCR technique have been efficiently applied in the identification of nitrifying population in aquatic and soil environments. Voytek and Ward (17) were one of the first to describe PCR primers specific for ammonium-oxidizer and successfully amplified nitrifier genes coding for rRNA from DNA extracts from natural samples. The PCR has been used to develop a sensitive and specific assay for detecting ammonium-oxidizing bacteria in the beta-subclass of the class *Proteobacteria*. PCR primers are selected on the basis of nucleic acid sequence data currently available for seven species of nitrifying bacteria in this subclass. The specificity of the ammonium oxidizer primers is evaluated by testing known strains of nitrifiers, several serotyped environmental nitrifier isolates, and other members of the *Proteobacteria*, including four closely related nonnitrifying species determined by rRNA sequence analysis.

Nitrifying populations from a series of soils and lakes have been differentiated by using DNA sequences of a ribosomal operon amplified by PCR, e.g., DNA extracts from 19 bacterioplankton samples collected from Antarctica and the Southern California Bight were assayed for the presence of ammonium oxidizers (18). This application of PCR is of particular importance for the detection and study of autotrophic nitrifiers, which are difficult to isolate from

indigenous microbial communities. 16s rDNA sequence information has been extensively exploited for the cultivation-independent analysis of ammonia oxidizers by PCR, quantitative dot-blot hybridization and denaturing gradient gel electrophoresis (19–21). These studies provide a sound understanding of the natural diversity of ammonia oxidizers.

Recently, FISH (fluorescent in situ hybridization) with oligonucleotide probes targeting signature regions of the 16s rRNA of ammonia oxidizers has been developed for the identification and quantification of nitrifying bacteria in natural and engineered systems. Based on the phylogenetic relationships of ammonia- and nitrite-oxidizing bacteria discovered by Head et al. (22), a 16s rRNA-targeted oligounucleotide probe for some chemolithoautotrophic ammonia-oxidizers was developed and successfully applied for the in situ detection of these bacteria in samples from sewage treatment plants (23–25). There is strong evidence that the FISH is a powerful molecular tool for rapid, reliable and cultivation-independent monitoring of phylogenetically defined bacterial populations in environmental samples (11, 25). 16S rRNA sequence analysis-based molecular methods offer the potential to better understand the composition and structure of biofilms (11, 26–28). The relevant percentage of nitrifiers and heterotrophs can be quantified by hybridization of ^{32}P or fluorescent-dye labeled oligonucleotide probes targeted to the unique regions of the 16S rRNA in them. This probe method is a useful tool for carrying out the sophisticated research needed to improve nitrification performance (25–28).

Other methods have also been applied for detecting nitrifying bacteria. Enzyme-linked immunosorbent assay (ELISA) using monoclonal antibodies (MAbs) is quite useful and helpful for the detection and quantification of nitrifying bacteria in a mixed microbial habitat. Inamori et al. (29) raised seven monoclonal antibodies from splenocytes of mice (BALB/c) that are specific for the surface antigen of the two kinds of nitrifying bacteria. The monoclonal antibody assay provides a quick and powerful means for estimating nitrifying population; the quantification of nitrifying bacteria could be completed within 5 hours. This method was used to rapidly enumerate the cells of *Nitrobacter sp*. in the influent to the Bromma Wastewater Treatment Plant (30).

In situ methodologies targeting the cbbL gene have been used to visualize cells of nitrifying bacteria. Prokaryotic in situ PCR (IS-PCR) and in situ reverse transcription (ISRT) protocols are currently employed to determine gene presence and expression, respectively (31). Aged-oligotropic seawater samples were inoculated with microbial assemblages containing a mixture of actively growing nitrifying bacteria, starved nitrifying bacteria, and heterotrophic bacteria without cbbL. After the molecular manipulations, all the nitrifiers (healthy or starved) with the cbbL gene were detected by IS-PCR, whereas only the actively growing autotrophic nitrifiers with detectable levels of carbon fixation and nitrification activity were detected by ISRT analysis (31). These results suggest that the combined IS-PCR and ISRT method has great potential for the analysis of heterogeneous populations where an assortment of healthy and starved/dormant cells are expected. It seems that more information of nitrifying bacteria could be obtained by combining different detecting methods. Daims et al. (32) analyzed the nitrifying diversity and population structure of a sequencing biofilm batch reactor receiving sewage with high ammonia and salt concentrations by the full-cycle rRNA approach, whereas the diversity and quantification of ammonia-oxidizers in the reactor

was additionally investigated using comparative sequence analysis of a gene fragment of the ammonia monooxygenase (amoA), which represents a key enzyme of all ammonia-oxidizers. On the other hand, the morphology of nitrite-oxidizing, unculturable *Nitrospira*-like bacteria was studied using FISH, confocal laser scanning microscopy (CLSM) and three-dimensional visualization, and microautoradiography combined with FISH. Rittmann et al. (33) not only assessed the community structure of activated sludge directly by oligonucleotide probes for seven treatment plants in France and two treatment plants in The Netherlands, but also employed mathematical modeling to translate the operating data into parameters that reflect community function and structure. These model-translated measures were further compared to the direct probing measures of community structure and more information was obtained from the correlation of these two measurements. Considering that the presence of ammonia oxidizers can be correlated with their characteristic activity, Wagner et al. (23) compared the probe-based enumeration with nitrification rates and provided a basis on which to measure the specific in situ activity per single cell.

The nitrification process consists of several oxidation reactions that involve various intermediates and intracellular compounds. To monitor the nitrogen compounds, researchers had turned their focus towards the development of stable and fast-responding sensors that could be placed directly in the liquor of the reactors. Based on bacterial reduction of the ionic species to N_2O gas, long-term stable biosensors for NO_x^- and NO_2^- had been developed, and the performance of those sensors were satisfactory in the testing period of several months in real wastewater treatment systems (34–36). Furthermore, because the nitrification process is closely associated with oxygen and other gaseous compounds (carbon dioxide, nitrogen etc.), an off-gas measurement approach has been adopted to develop a respirometer based on gas phase mass balance with a mass spectrometer as off-gas detector (37). This respirometer can be used to investigate the effects of dissolved oxygen, carbon source and pH on nitrification/denitrification. In addition to the traditional MPN and fluorescent antibody methods, respirometric techniques can improve chemical analyses of intermediates and intracellular compounds produced in nitrification and denitrification processes. Indeed, combined gene-based identification, microbial physiology techniques and biosensors would provide more information for monitoring, tracking and understanding of nitrifying bacteria, which are slow-growing, difficult culturing, but essential for nitrogen transformation in the environment.

2.4. Nitrification Kinetics

Nitrification kinetics examines the factors that can affect the rates of nitrification reactions as well as the microbial synthesis of nitrifying bacteria. The growth of either ammonium oxidizer or nitrite oxidizer is limited by the concentration of ammonium and nitrite, respectively. According to the best-known Monod equation, the growth kinetics of nitrifying bacteria can be written in Equation (7):

$$\mu_n = \mu_{n,max} \frac{S}{K_s + S} \tag{7}$$

K_s values for both ammonium oxidizer and nitrite oxidizer are around 1 mg N/L at temperatures below 20°C, besides, the rate-limiting step for nitrification in municipal wastewater

treatment systems seems to be the first-stage oxidation of ammonia by ammonium oxidizer because nitrite has rarely been observed to accumulate, except in the presence of substances that are selectively inhibiting to nitrite oxidizer (38). According to the general relationships among the specific substrate removal rate, specific growth rate and growth yield coefficient, the relationship between the oxidation rate of ammonium and the growth rate of ammonium oxidizer can thus be described in Equation (8):

$$q_N = \frac{\mu_N}{Y_N} = q_{N,max}\frac{[NH_4\text{-}N]}{K_N + [NH_4\text{-}N]} \tag{8}$$

The growth of microorganisms may be expressed in terms of their doubling or generation time. It should be noted that the generation times of nitrifying bacteria are 10 to 20 times longer than those of heterotrophic bacteria. To retain an adequate population and amount of nitrifying bacteria, a long solids retention time (also referred to as the mean cell residence time or sludge age) is required in the nitrification process. The solids retention time (SRT) in a biological system is normally defined as the total mass of biological solids present in the system over the total mass of biological solids leaving the system daily. At steady state, the solids leaving the system will be equal to the solids produced. Therefore, the growth rate and SRT of the organisms in the system are related by:

$$\frac{1}{SRT} = \mu_n - K_d \approx \mu_n' \tag{9}$$

For nitrifying bacteria, K_d is often negligible. In this case, the specific growth rate μ_n is the same as the net specific growth rate μ_n'.

2.5. Factors Affecting Nitrification

The growth of nitrifying bacteria is very sensitive to changes in culture environments. Intensive research efforts have been focused on the factors that could influence nitrification kinetics and growth kinetics of nitrifying bacteria. These factors include substrate availability, temperature, pH, DO concentration and presence of inhibitors.

2.5.1. Substrate Availability

As discussed earlier, nitrification can be completed by two kinds of bacteria: ammonium and nitrite oxidizers. Ammonium is converted to nitrite by ammonium oxidizer, and then to nitrate by a nitrite oxidizer. There is evidence that both ammonium and nitrite oxidizers are sensitive to high concentrations of their own substrates, and more so to the substrate of each other (39). To date, it has been recognized that free ammonia (FA) can inhibit both ammonium and nitrite oxidizers, whereas nitrite oxidizer is much more sensitive to FA than ammonium oxidizer. The FA concentration in reactor can be estimated using the expression proposed by Ford et al. (40):

$$FA = \frac{[NH_4\text{-}N] \times 10^{pH}}{\exp[6334/(273 + T)] + 10^{pH}} \tag{10}$$

This equation shows that the *FA* concentration is strongly dependent upon the culture pH. Inhibition of nitrification by free ammonia and free nitrous acid has been described in

many literatures. It is accepted that the respective *FA* inhibition threshold is 10 to 150 mg/L for ammonium oxidizer, and 0.1 to 4.0 mg/L for nitrite oxidizer (41–43). Lower inhibition threshold of free ammonia to nitrite oxidizer would lead to the accumulation of nitrite in wastewater containing high strength ammonium. Nitrite in turn may promote the production of nitrous acid, which further inhibits nitrite oxidizer in the range of 0.2 to 2.8 mg/L.

2.5.2. pH and Alkalinity Effects

Hydrogen ions produced during the nitrification process would result in substantial destruction of culture alkalinity. Evidence shows that the optimal pH for nitrification falls into a very narrow range of 7.8 to 8.0, and the culture pH higher than the optimal values would inhibit the activity of nitrifying bacteria owing to the pH-enhanced production of free ammonia. As the pH moves to the acid range, the rate of ammonium oxidation declines, and this tendency is true for unacclimated as well as for acclimated cultures, however the acclimation or selection of different populations of nitrifying bacteria with culture time, in some extent, can moderate pH effects. The rate of nitrification and the number of nitrifying bacteria would decline to pH values below 6.0, whereas nitrification would be completely blocked at pH values lower than 5.0.

In a biofilm reactor, nitrification efficiency was reduced by 50% at pH 6.0 after 1.5 days of acclimation, but no decline in nitrification efficiency was found after 10-day acclimation (44). No adverse effect on nitrification was observed when the reactor pH was abruptly reduced from 7.2 to 6.4, whereas nitrification was partially repressed when the pH was lowered from 7.2 to 5.8, indicated by a sharp increase in the effluent ammonium concentration from approximately zero to 11 mg N/L. However, when the culture was restored to the initial pH of 7.2, nitrifying bacteria could rapidly recover their microbial activity, this implies that the lower pH was only inhibitory, but not toxic to nitrifying bacteria (45, 46). Consequently, for design purposes, it is sufficient to consider that the nitrification rate may drop significantly as the pH is lowered below the neutral range, and that for performance stability, it is best to maintain a pH at 6.5 to 8.0.

2.5.3. Temperature

Nitrifying bacteria can survive over a very wide range of environmental temperature, from 4 to 45°C (5). There is convincing evidence that the maximum growth rate and the half-saturation coefficient of nitrifying bacteria are temperature-sensitive, and some Arrhenius-type expressions for the effect of temperature on the maximum growth rate and half-saturation coefficient of ammonium oxidizer over a temperature range of 5–30°C was proposed suspended growth process design (39):

$$\mu_{N,max} = 0.18e^{0.116(T-15)} \tag{11}$$

$$K_n = 0.405e^{0.118(T-15)} \tag{12}$$

2.5.4. Dissolved Oxygen

Nitrification is an obligatorily aerobic process, i.e., molecular oxygen is final electron acceptor. For pure cultures of ammonium and nitrite oxidizers, the critical dissolved oxygen concentration below which nitrification does not occur is around 0.2 mg/L (47). However, in

full-scale wastewater treatment process, for suspended nitrification to proceed, the dissolved oxygen level should not be less than 2 mg/L, whereas the optimal maximum DO concentration (at the end of the aeration period) for nitrogen removal was determined to be 2.0 to 2.5 mg/L (48, 49). As compared to suspended growth system, the dissolved oxygen concentration in attached growth reactors should be relatively high; the bulk fluid dissolved oxygen levels should be near 70% saturation to prevent oxygen-transfer limitations inside biofilms. In real wastewater treatment systems, the effects of ammonia, temperature, pH, and dissolved oxygen on the nitrification process are interrelated. It had been proposed that the combined effect of these factors on biological nitrification can be described in the form of the Monod-type expression (39, 50):

$$\mu_n = \mu_{n,max} \times \frac{[NH_4\text{-N}]}{K_n + [NH_4\text{-N}]} \times \frac{DO}{K_o + DO} \times [1 - 0.833(7.2 - pH)] \tag{13}$$

2.5.5. Effect of C/N Ratio

The ratio of the feed biodegradable organic carbon to the nitrogen available for nitrification in the wastewater, namely the C/N, is one of the critical factors associated with the performance of nitrification systems. Okabe et al. (51) investigated the effects of different C/N ratios on time-dependent population dynamics of nitrifiers and heterotrophs in undefined mixed-population biofilms as well as on nitrification efficiency. The results showed that the population dynamics and nitrification efficiency were strongly related to the initial microbial composition in the biofilms and C/N ratio. It seems that a higher C/N ratio would retard the accumulation of nitrifying bacteria, especially NO_2-oxidizers. These in turn result in a considerably long start-up period for complete and stable nitrification owing to competition for dissolved oxygen and space in the biofilms. Because the growth yield of heterotrophic bacteria is much greater than that of nitrifying bacteria, organic matter in the reactor could sustain a faster growth of heterotrophic bacteria. If the growth rate of heterotrophic bacteria established exceeds the maximum growth rate of nitrifying bacteria, nitrifying bacteria would be washed from the treatment system. For biological treatment process incorporating a nitrification unit, the overall solids retention time should be designed to be greater than the minimum solids retention time required for nitrification at environmental conditions within the reactor to build and sustain nitrifying population.

2.5.6. Inhibitory Effect

Nitrifying bacteria are slow-growing organisms, and they are particularly susceptible to inhibitory and toxic compounds. A wide variety of organic and inorganic chemicals can repress the growth of nitrifying bacteria, and an ammonium oxidizer is more susceptible than nitrite oxidizer. Most heavy metals are inhibitory to nitrifying bacteria. Sato et al. (52) found that the inhibition of copper and nickel to *Nitrosomonas europaea* was highly correlated to the amine compounds of copper and nickel, while research by Lee et al. (53) showed that nitrifying bacteria were more sensitive to copper than nickel, and *Nitrosomonas* sp. was more sensitive to copper and nickel than *Nitrobacter* sp. Based on measurements of ammonium uptake rate (AUR) and specific oxygen uptake rate (SOUR), Paolo et al. (54) concluded that

for activated sludge taken from a full-scale nitrifying plant, the inhibitory order of heavy metals was Cd > Cu > Zn and Pb > Cr.

2.5.7. Influence of Oxidative-Reductive Environments

To remove carbon together with the removal of nitrogen and phosphorus, some systems run alternatively under aerobic, anoxic and anaerobic conditions. Operational practice shows that anoxic or anaerobic conditions lasting for hours have no tangible impact on nitrifier viability when acceptable dissolved oxygen concentrations are restored.

3. FUNDAMENTALS OF DENITRIFICATION PROCESS

Biological denitrification is the microbial reduction of nitrate to nitrite, and ultimately to nitrous oxide and/or nitrogen gas. Nitrate and nitrite are electron acceptors instead of molecular oxygen for microbial respiration in this process. Thus, denitrification is commonly thought to occur only in the conditions without the presence of molecular oxygen that is referred to as anoxic.

3.1. Microbiology

Unlike nitrification, a relatively broad range of bacteria can accomplish denitrification. Denitrifiers are ubiquitous in most natural environments, including municipal wastewaters and sludges (55). This is partially owing to the fact that denitrifying bacteria are facultative, i.e., they can use either oxygen or nitrate as their terminal electron acceptor. In fact, denitrifying bacteria can proliferate in aerobic systems because of their ability to use oxygen and efficiently oxidize organic matter (56). These characteristics of denitrifying bacteria indeed minimize the need to create special environmental conditions for their survival as compared to nitrifying bacteria.

Denitrifying bacteria are a biochemically and taxonomically diverse microbial group, and they may be organotrophs, lithotrophs, phototrophs, diazotrophs and so on. Although some denitrifying bacteria are chemoautotrophs that can use hydrogen or reduced sulfur compounds as energy sources, and others are photoautotrophs, most of these organisms generally derive their energy from the oxidation of fixed carbon substrates, including single carbon compounds. The primary substrate and end product of denitrification process are nitrate and nitrogen gas, but some denitrifiers can reduce nitrite only, namely nitrite-dependent denitrifiers, whereas others lack nitrous oxide reductase, thereby produce nitrous oxide as the terminal product. It is unrealistic to group denitrifying bacteria into one or a few bacterial genera owing to the variety of metabolic types of bacteria capable of denitrifying. Denitrifying bacteria must meet the following criteria. (i) At least 80% of the nitrate or nitrite reduced by the bacterium must be converted to nitrogen gas and nitrous oxide; (ii) There must be an increased growth yield owing to the reduction of nitrate, nitrite or nitrous oxide. This trait is the primary requirement for classing an organism as a denitrifier; (iii) The conversion of nitrate to nitrous oxide and nitrogen gas must occur at a high rate, i.e., the process must be central to cellular intermediary metabolism, not just a side reaction providing a minor pathway for electron transport; (iv) The presence of cytochrome cd or disimilatory nitrite reductase should be demonstrable in the microbial cells (57). Therefore, caution should be exercised when classification of

denitrifying bacteria is strictly based on the conversion of nitrate or nitrite to nitrous oxide or nitrogen gas.

3.2. Stoichiometry

The denitrification process involves the transfer of electrons from electron donor (i.e.,, carbon substrate) to the electron acceptor (i.e., oxygen, nitrate or nitrite). For the electron donor, it could be either the organic substrate in the raw wastewater or a substrate added to the source. The most commonly used external carbon source is methanol when dentrification is accomplished as a separate stage. The reaction of denitrification can be generally expressed as follows (39):

$$NO_3^- + 5/6\,CH_3OH \rightarrow 5/6\,CO_2 + 1/2\,N_2 + 7/6\,H_2O + OH^- \tag{14}$$

Based on the above equation, the overall cell synthesis in denitrification process may be written as (58):

$$NO_3^- + 1.08\,CH_3OH + 0.24\,H_2CO_3 \rightarrow 0.056\,C_5H_7NO_2 + 0.47\,N_2 + 1.68\,H_2O + HCO_3^- \tag{15}$$

Although the distribution of the organic substrate between incorporation into new biomass versus that used in the reduction of nitrate is a function of organic compounds used, microbial population, and the operating conditions, the above equations give the general ratio about the requirement of organic carbon and nitrogen in the denitrification process.

3.3. Metabolisms

Denitrifying bacteria may use the route of denitrification as an alternative to normal aerobic respiration. When oxygen is available, aerobic respiration is a main metabolism of denitrifying bacteria. The synthesis of nitrate reductase is repressed by oxygen, whereas nitrate and nitrite may act as electron acceptors in the respiratory electron transport chain after depletion of oxygen. Both processes are accomplished through use of cytochromes in an electron transport chain (59). However, in cases where oxygen is present, different cytochromes are needed for the reduction of oxygen to water, and dissimilatory nitrate reduction is inhibited (60).

According to the chemiosmotic mechanism, for most of aerobic bacteria, ATP is generated by oxidative phosphorylation. In this process, electrons are transported through the electron transport system from an electron donor (substrate) to a final electron acceptor (O_2). Molecules directly using the H^+ gradient built up by electron transport can be considered as H^+-ATP pumps. In methanogens, ATP synthesis is linked with methanogenesis by electron transport, proton pumping and a chemiosmotic mechanism (5). Similar to aerobic respiration, anaerobic respiration is useful because it is more efficient than fermentation and allows ATP synthesis by electron transport and oxidative phosphorylation in the absence of oxygen. Thus, it seems that the proton translocation-driven phosphorylation would be a common mechanism for energy generation in both aerobic and anaerobic respirations.

It must be pointed out that some bacteria, e.g., *Streptococcu*s, have no respiration chain and can produce ATP only via substrate-level phosphorylation. The electron transport chain is the fundamental mechanism by which cells generate energy. The process involves transferring

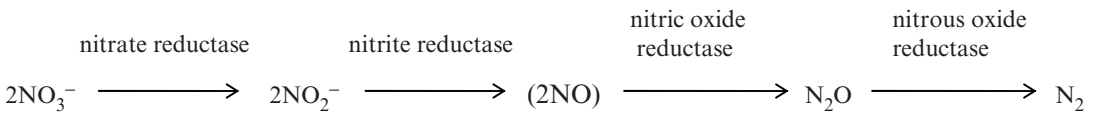


Fig. 13.3. Series of reduction reactions in denitrification.

electrons from a reduced electron donor (e.g., an organic substrate) to an oxidized electron acceptor (e.g., oxygen, nitrate, nitrite, or sulfate). Nitrate or nitrite may serve as a substitute of oxygen in the respiration chain with only small modifications to the metabolic systems (i.e., the enzymes) of the bacteria. By using nitrate or nitrite in place of oxygen in the electron transport chain, however, less energy is generated. This is owing to the fact that these alternate electron acceptors have less positive reduction potentials than molecular oxygen (5). Energy yield is directly related to the magnitude of the reduction potential difference, thus less energy available to make ATP is generated from denitrification.

During biological denitrification, nitrate is transformed to nitrogen gas by a series of reduction reactions (Figure 13.3). Nitric oxide is generally enclosed in brackets in the depiction of this reaction sequence because it is not usually detected as a free intermediate. It is still uncertain whether nitric oxide is a true intermediate in the process or a side reaction (61, 62). It is believed increasingly that nitrate is reduced to nitrous oxide and nitrogen gas with the possible transient accumulation of nitrite plus the occasionally detected transient accumulation of nitric oxide (63, 64).

The four-step denitrification as shown in Figure 13.3 is catalyzed by a series of enzymes. Reduction of nitrate to nitrite is catalyzed by the nitrate reductase, which is a membrane-bound protein. Both the synthesis and the activity of nitrate reductase can be inhibited by the presence of molecular oxygen. The second enzyme in this pathway is nitrite reductase, which catalyzes the conversion of nitrite to nitric oxide. Synthesis of nitrite reductase is induced by nitrate, but can be repressed by oxygen. Nitric oxide reductase, a membrane–bound protein, is responsible for the conversion of nitric oxide to nitrous oxide. The synthesis of this enzyme is inhibited by oxygen and induced by various nitrogen oxide forms. In the last step, nitrous oxide is converted to nitrogen gas by nitrous oxide reductase. This is a periplasmic copper-containing protein, which can be inhibited by low pH and is much more sensitive to oxygen than the other three enzymes in the denitrification pathway. This is the reason why nitrous oxide instead of nitrogen gas would be the final product of denitrification when high oxygen and low pH prevail in cultures (65).

In pure cultures, oxygen has been found to repress the synthesis and further activity of the enzymes required for denitrification; however the synthesis of these denitrifying enzymes can be induced when oxygen is depleted in the culture in which denitrifying bacteria reside. After the control mechanisms in denitrifying bacteria switch from aerobic to anoxic conditions, development of a maximally functional denitrification system requires about 40 minutes to three hours (66). Also evidence from the activated sludge process indicates that the denitrifying enzymes may be present even in systems that do not have anoxic conditions, suggesting

that a period for synthesis of the denitrifying enzymes is not strictly required before initiation of denitrification (67). In summary, both the synthesis and activity of denitrifying enzymes are controlled by the availability of molecular oxygen. Comparatively, the oxygen inhibition to the activity of nitrifying enzymes is more significant than to the enzyme synthesis. The activity of *Paracoccus denitrifica,* is restricted to anaerobic conditions. On a change from aerobic to anaerobic respiration, a culture of *P. denitrificans* enters an unstable transition phase during which the denitrification pathway is induced, however after switching back to aerobic condition, denitrification of the cells stops at once, although sufficient nitrite reductase is still present (67).

Because many intermediates, e.g., HNO_2, NO, and N_2O, would be generated from the denitrification process, one engineering concern is the release of these intermediates into the environment. It is clear that these intermediates can be readily formed under electron donor limitation. When cells are subject to transitions between aerobic and anoxic conditions, the formation of these intermediates was found to be enhanced (68, 69). This might be owing to the enzyme regulation inside the cell that cannot respond immediately to a changed environment (67, 70). An important but regularly neglected aspect concerning the possible emission of denitrification-associated N_2O into the environment is its high solubility in water. N_2O is not easily stripped into air and an initial accumulation can be followed by consumption when the bacteria have been adapted for the change from aerobic to anoxic conditions in several minutes. Because stripping in large-scale installations is an order of magnitude less than in laboratory scale units, extrapolation of lab studies might easily lead to an overestimation of the emissions of N_2O (71).

3.4. Methods for Identifying Denitrifiers

The population density of denitrifiers is commonly estimated through use of most-probable-number (MPN) procedures, but the MPN method is inherently imprecise and time-consuming (72). This method is based on detecting the conversion of nitrate or nitrite to gaseous end products in liquid culture media (Figure 13.2). More accurate and sensitive methods for enumeration of denitrifying bacteria have been well-established along with the development of molecular microbiological techniques (73). Similar to the identification of nitrifying bacteria, denitrifying bacteria can also be studied by the methods of nucleic acid probes, e.g., DNA probe and RNA probes, PCR technique, fluorescent in situ hybridization (FISH), confocal laser scanning microscopy (CLSM), three-dimensional visualization, enzyme-linked immunosorbent assay (ELISA) separately or in combination (74, 75). In addition to the above methods, flow cytometry (FCM) method has been used to detect denitrifying bacteria. To examine the abundance of the denitrifier *Pseudomonas* sp. JR12, Tal et al. (76) used immuno-labeling of the *Pseudomonas* followed by FCM to determine the relative abundance of this bacterium under the various incubation conditions. The results of FCM were further compared with the results obtained from the ELISA, and the close agreement between these two detection methods was established (76). Therefore, it seems that FCM could be a rapid and accurate tool for the detection of the relative abundance of immuno-labeled target organisms in heterogeneous microbial populations.

3.5. Procedures for Measuring Denitrification

Tracer N, mass spectrometry, emission spectrometry, gas chromatography, and the acetylene blockage technique are commonly used to determine the extent of denitrification in aquatic systems. Each of these approaches has its own advantages and disadvantages. Procedures for mass and emission spectrometry are presented in "Nitrogen Isotope Techniques", in which sample handling, instrumentation, and data calculations are detailed for all the main phases of the N cycle, including denitrification (77, 78). Procedures for measuring nitric oxides and rapid isotopic analyses at atmospheric concentrations are also available (79, 80). Because nitrous oxide reductase can be inhibited by acetylene, blocking the denitrification sequence at the N_2O stage offers a relatively simple approach for studying the denitrification process. The acetylene blockage has been extensively exploited in defining the parameters of denitrification. Measuring small amounts of evolved N_2 against the large background of atmospheric N_2 requires use of ^{15}N, whereas N_2O is easily measured by gas chromatography. It should be emphasized that N_2O is also produced during nitrification, and the N_2O produced during nitrification slightly exceeds that released from denitrification.

3.6. Denitrification Kinetics

The specific growth rate of denitrifying bacteria can be related to the concentration of nitrate by a Monod-type expression such that

$$\mu_D = \mu_{D\,max} \frac{S_{NO3}}{K_D + S_{NO3}} \tag{16}$$

This equation reduces to zero-order kinetics if K_D is considerably less than S_{NO3}, while a first-order expression if S_{NO3} is significantly less than K_D. In the environmental engineering, the net specific growth rate (μ') of microorganisms in a system is the inverse of *SRT*:

$$\frac{1}{SRT} = \mu'_D \tag{17}$$

The growth rate of denitrifiers is in the same level as aerobic heterotrophic organisms and much greater than nitrifiers. Thus, the minimum SRT required to prevent the washout of denitrifying bacteria from a reactor is much shorter than that designed for the nitrification process. It is known that the denitrification process is strongly dependent upon the concentrations of organic substrate and dissolved oxygen present in the treatment system. Analogous to the Monod equation, a multi-parameter kinetic model has been proposed for the rate of nitrate removal in denitrification process (81),

$$q_{NO3} = q_{NO3,max} \frac{S_{NO3}}{K_{NO3} + S_{NO3}} \times \frac{S_C}{K_C + S_C} \times \frac{1}{K_o + DO} \tag{18}$$

3.7. Factors Influencing Denitrification

The main factors affecting denitrification processes are as follows: the nature and amount of organic matter, nitrate concentration, aeration status, pH and temperature.

3.7.1. Availability of Carbon and Nitrate

Denitrification is mainly accomplished by heterotrophic bacteria, and is strongly dependent on the availability of organic carbon, which serves as an energy source and electron donor of the denitrification process. A wide variety of organic carbon can fulfill the requirements of denitrification. In designing of full-scale denitrification process, the selection of organic carbon depends upon three factors: availability of the organic carbon, the reaction rate, and costs. In industry practice, methanol has been usually used as external carbon source for denitrification because of its low cost and achievable high denitrification rate. Other potential sources of organic carbon, including the organics present in wastewater, are described in the environmental engineering literature. Hallin and Pell (82) studied changes in the functional properties of denitrifying bacteria adapted to methanol and ethanol, respectively. Short-chain fatty acids were less popular for denitrifiers in the methanol-adapted sludge than they were in the reference sludge. Denitrification capacity with ethanol and acetate in the ethanol-adapting sludge increased rapidly, and the ethanol-adapted sludge also had a higher capacity to denitrify with butyrate, glycerol and methanol. Denitrification rates with wastewater organics are approximately one-third of those in methanol-denitrification systems (4). One experiment attempted to use inorganic compounds as electron donors, such as hydrogen and sodium sulphide (83). There is evidence that at low nitrate concentrations, nitrate reduction is subject to be a first-order kinetics; however, for nitrate concentration greater than 20 mg N/L, the denitrification reaction seems to be a zero-order kinetics, independent of the amount of nitrate present (84). Denitrification is an enzyme-catalyzed biological process, which generally follows the Michaelis-Menten equation. Beyond the enzyme-saturating concentrations of nitrate, nitrate would be inhibitory to denitrifying bacteria, i.e., a high nitrate concentration may inhibit the activity of nitrogen oxide reductases (57).

The value of half-saturation constant of denitrification is very low. This implies that the denitrification process can be run at quasi-maximum unit removal rates, and the denitrification rate may be determined by the amount of organic carbon available for metabolism rather than by the nitrate level. Nitrate concentration with respect to organic carbon supply can influence the N_2O/N_2 ratio. Hiroki et al. (85) investigated the effects of influent COD/N ratios on N_2O emission from a biological nitrogen removal process with intermittent aeration, and fed with high-strength wastewater. It was found that in steady-state operation, 20–30% of influent nitrogen was emitted as N_2O in the bioreactors with influent COD/N ratios less than 3.5, and high N_2O emission rates at low COD/N ratios was mainly owing to endogenous denitrification.

3.7.2. Aeration

As discussed earlier, the presence of high dissolved oxygen concentration may repress the synthesis of a series of reductase involved in denitrification. For practical purposes, denitrification can be ignored when dissolved oxygen concentrations are greater than 1.0 mg/L (86).

3.7.3. pH

Denitrification may occur within the pH range of 3.9 to 9.0, and the maximum nitrogen oxide reduction rate falls into pH 7.0 to 8.0 (85). Most denitrifying bacteria grow best at

pH 6 to 8, but denitrification could be hindered and still remain significant below pH 5. Denitrification by organotrophs is negligible or absent when the culture pH drops below pH 4.

3.7.4. Temperature

The effect of temperature on biological denitrification is often described by an Arrenhius-type function (4, 87):

$$q_{D,T} = q_{D,20} \times \theta^{(T-20)} \tag{19}$$

This equation shows the temperature-dependent effect on the activity of denitrifying bacteria, however, the temperature-related oxygen solubility and diffusivity in water should also be taken into account in a real process operation. Denitrification can occur in a wide range of culture temperatures, from 5 to 75°C (84). At temperature above 50°C, chemical decomposition of NO may become significant.

3.7.5. Inhibitory Effects

In general, the sensitivities of denitrifying and aerobic heterotrophic bacteria to inhibitory compounds are comparable. Thus, commonly applied chemical concentrations that result in inhibition to heterotrophic respiration can be used as important references for denitrification. The US EPA published data of chemical inhibition thresholds for activated sludge and trickling filters (88). When reviewing these threshold data, the ability of a biomass to acclimate to high levels of inhibitory compounds should be taken into account because the acclimated cultures can tolerate much higher concentrations of inhibitory compound than unacclimated systems do.

4. MODELING OF NITRIFICATION AND DENITRIFICATION

Mathematical modeling is a technique that has been increasingly used to analyze problems of significance to environmental engineering. A wide variety of models are available to characterize full-scale nitrification and denitrification systems. The models for suspended growth systems are the most fully developed and tested, whereas for fixed-film processes, ongoing research has directed to the development of improved models.

4.1. Suspended-Growth Models

The IAWQ Task Group on Mathematical Modeling of Activated Sludge processes proposed Activated Sludge Model No.1 in 1987 (89). This kinetic model describes the dynamic behavior of systems incorporating COD removal and N removal by nitrification-denitrification processes. Through its successful use and application, the model has achieved widespread acceptance and has had a significant impact on the approach to design, operation and control of nitrification-denitrification systems. The revised version, Activated Sludge Model No. 2, has been recently published (90). The No. 2 model provides a powerful mathematic tool for the simulation of combined biological phosphate, organic carbon and nitrogen removal process.

4.2. Fixed-Growth Models

The modeling of fixed growth systems is more difficult than that of suspended growth systems. This is probably owing to the effect of diffusional resistance on the substrate removal rate that must be incorporated into the biofilm model, and the heterogeneous and poorly characterized conditions that occur within many fixed growth reactors. In spite of these difficulties, research continues on the development of models of fixed-film processes. Generally, biofilm models are set up based on the calibration and optimization of Activated Sludge Model No. 1 and No. 2, according to specific data of the nitrification/denitrification systems.

5. BIOLOGICAL NITRIFICATION AND DENITRIFICATION PROCESSES

Owing to advantages such as high removal efficiency, process stability/reliability; relatively easy process control; low land area requirements and moderate cost, biological nitrification/denitrification becomes an increasingly attractive method for the removal of nitrogen. It is a two-step process in which ammonia is converted aerobically to nitrite/nitrate (nitrification), then the nitrite/nitrate are reduced to nitrogen gas (denitrification). Almost all operating biological nitrogen removal processes can be classified into three groups (Figure 13.4). According to this classification, the biological nitrification processes include two configurations, such as combined carbon oxidation/nitrification and separate-stage nitrification, whereas biological denitrification processes are categorized as combined carbon oxidation/nitrification/denitrification (single-sludge denitrification) and separate-stage denitrification.

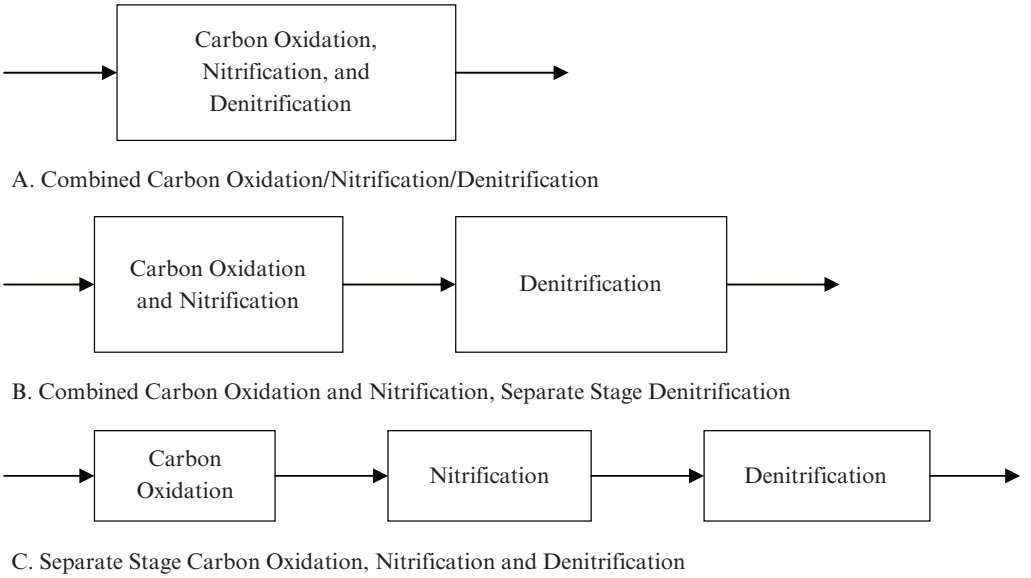


Fig. 13.4. Three major approaches to biological nitrogen removal.

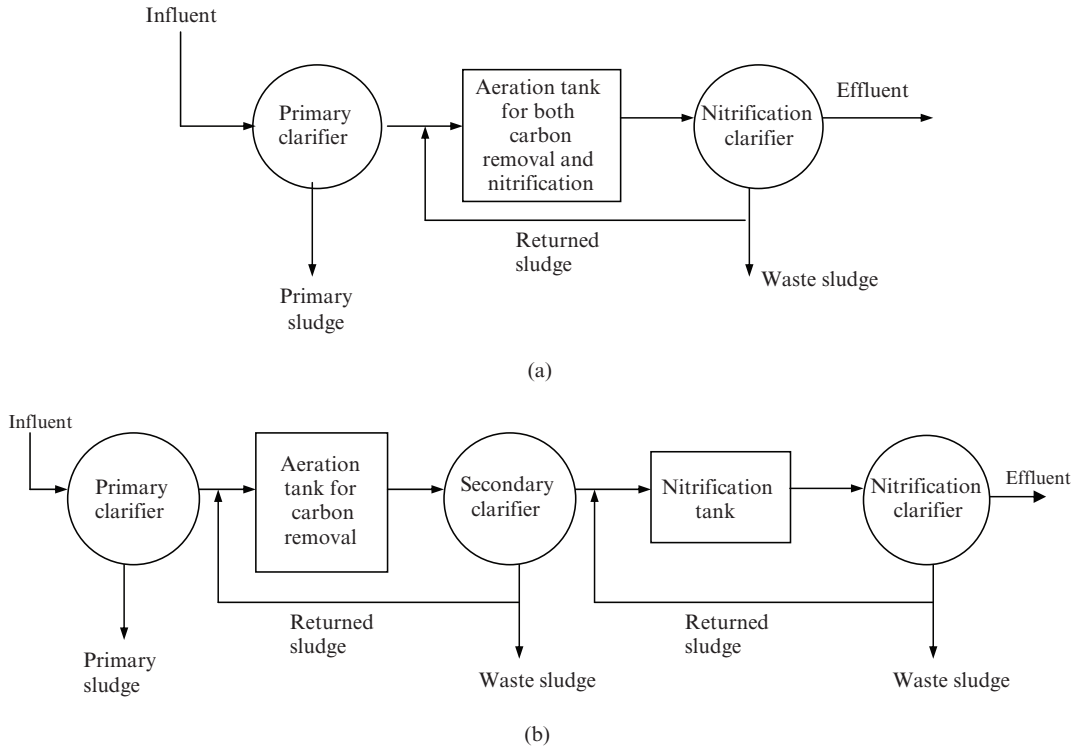


Fig. 13.5. Typical suspended-growth carbon oxidation and nitrification processes: (a) single-stage and (b) separate-stage.

5.1. Nitrification Processes

Biological nitrification process in wastewater treatment is particularly applicable to cases in which only ammonia removal is required, but also is a key step towards biological nitrogen removal. The combined carbon oxidation and nitrification may occur in a single reactor, termed "single stage." In separate-stage nitrification, carbon oxidation and nitrification occur in separated units. Suspended- or attached-growth reactors may be used for either single-stage or separate-stage systems. Examples of single-stage and separate-stage nitrification are illustrated in Figure 13.5.

5.1.1. Combined Carbon Oxidation/Nitrification

Combined carbon oxidation/nitrification process can be subdivided into suspended-growth process, attached-growth process and combined suspended-and attached-growth systems.

5.1.1.1. Suspended-Growth Processes

Nitrification can be accomplished in any of the suspended-growth processes, including conventional plug flow, complete-mix, extended aeration and various modification of the oxidation ditch as presented in Table 13.1. Successful nitrification relies on several operation

Table 13.1
Main biological treatment processes used for nitrification

Type	Common name	Use
Aerobic Processes:		
Suspended-growth	Activated-sludge process	Carbon removal (nitrification)
	Conventional (plug-flow)	
	Step aeration	
	Complete-mix	
	Sequencing batch reactor	
	Pure oxygen	
	Contact stabilization	
	Extended aeration	
	Deep shaft	
	Deep tank	
	Oxidation ditch	
	Aerated lagoons	Carbon removal (nitrification)
Attached-growth	Tricking filters	Carbon removal (nitrification)
	Rotating biological contactors	Carbon removal (nitrification)
	Packed-bed reactors	Carbon removal (nitrification)
	Fluidized-bed rectors	Carbon removal (nitrification)
	Biological aerated filters	Carbon removal (nitrification)
Combined suspended and attached-growth processes	Activated biofilter process, tickling-filter solids-contact process, biofilter activated-sludge process, series trickling-filter activated-sludge process	Carbon removal (nitrification)
Combined aerobic, anoxic and anaerobic processes:		
Suspended-growth	Single- or multi-stage process, Various proprietary processes	Carbon removal, nitrification, denitrification
Combined suspended and attached-growth	Single- or multi-stage processes	Carbon removal, nitrification, denitrification

Adapted from (143).

parameters, such as ammonium, nitrite concentrations, BOD_5/TKN ratio, dissolved oxygen concentration, temperature, pH and sustainable growth of nitrifying bacteria and so on. The nitrification ability of the conventional activated sludge process is highly temperature-dependent. At low winter temperatures, nitrification can be sustained only if the activated sludge process is operated at relatively long solids retention time (SRT), otherwise the wash-out of nitrifying bacteria from the system would occur and cause nitrification to cease. To solve this problem, the short SRT nitrification process was developed by Kos (91) to provide an inexpensive alternative for plants that need to upgrade their activated sludge process for year-round nitrification or nitrogen removal. In this process, supplemental nitrifying bacteria that were cultivated in a separate small side-stream aeration tank using ammonia available in the digested sludge dewatering liquid or in the digester supernatant, were added daily to the main activated sludge unit to replenish nitrifying bacteria. The bioaugmentation-based strategy allows the main stream activated sludge process to efficiently nitrify during low winter temperatures at a very short SRT that provides nitrification in an aeration tank substantially smaller than required for conventional nitrification.

In case of nitrogen shock loads and/or toxicity incidents, spare nitrification capacity is usually needed for a nitrifying activated sludge plant. To avoid the traditional, over-designed plants with longer SRT, one approach is to store the spare biomass in a separate sludge storage tank, then return the stored biomass back to the main stream process when a shock nitrogen load or an inhibition/toxicity incident occurs (92). Following this strategy, the aeration tank volume could be reduced by 20%, as compared to the conventional system.

5.1.1.2. Attached-Growth Processes

Several attached-growth reactor configurations including trickling filters, rotating biological contactors, aerated biological filters, packed and fluidized-bed systems have been used to promote nitrification (Table 13.1). Novel attached-growth processes for combined organic carbon oxidation/nitrification appear in the environmental engineering literature. In the past few years, biological aerated filter (BAF) has attracted intense research attention. BAF in fact is a large-granule fixed-bed filter that is specially developed for secondary biological wastewater treatment. This system has the advantage of separating biodegradation and solids in a single stage. This eliminates the need for secondary settling or tertiary filtration, and the organic loading rates applied to the BAF could be 5 to 30-fold higher than those in the conventional biological processes. With a long SRT, the BAF promotes the accumulation of slow-growing nitrifying bacteria. Consequently, high concentration of nitrifying bacteria can be achieved and maintained through microbial attachment on the media. It has been demonstrated in full-scale BAF processes that the efficiency of combined BOD removal and nitrification as well as process stability are quite satisfactory (93, 94). However, bacteria could form thick biofilms on the carrier materials in BAF, leading to reduced diffusion efficiency of substrate/oxygen and unpredictable slough of attached biomass (95).

5.1.1.3. Combined Suspended-and Attached-Growth Systems

A variety of process configurations have been developed to combine suspended- and attached-growth components into a single treatment process (Table 13.1). These approaches

offer protection against biomass washout, improved handling of industrial discharges or toxic shock loads, improved SVI and SS settling velocities, and overall ease of operation. For the combined suspended- and attached-growth systems, more emphasis has been put on the development of novel media on which biofilms may form. Some new media have been trailed, such as freely moving porous pads used in the Captor and Linpor systems, trickling filter media (FAST and Bio-2-sludge), racks of open weave media fixed in place (Ringlace), and modular media systems (Monitor). The Captor and Linpor systems developed by Ashbrook-Simon-Hareley and Linde AG respectively, use porous pads freely suspended in the aeration basin. Linpor sponges are approximately cubical with sides of about 10 to 12 mm, and Captor sponges are about 12 by 25 by 25 mm (4). Either system may include but not require recycling of settled solids from the final settler back through the pads. The recycle option leads to a combined attached/suspended growth system. Another combined system, namely the Downflow Hanging Sponge-Cubes (DHS) reactor, consists of several hanging sponge-cubes strings composed of 90 polyurethane sponge-cubes connected diagonally in series with each other (96). The influent stream gravitationally migrates downward from the anterior cube to the posterior cubes toward the outlet. The DHS reactor is used as a post-treatment unit to the UASB effluent *in vivo*, and exhibits an excellent removal of the remaining COD of the UASB effluent as well as relatively high nitrification with an efficiency of 60% to 70%.

Submerged trickling filters include the FAST system by Smith and Loveless, Bio-2-Sludge by Weber Engineering. In a FAST system, typically 75% of the aeration tank volume is occupied by media. The media is completely submerged in water, and diffused aeration forces wastewater to flow up through the media. In contrast to the FAST systems, only about 25% of the aeration tank is occupied by the trickling filter media in the Bio-2-sludge system. In both processes, fixed-film and freely suspended biomass resulting from sludge recycle coexist.

In a Ringlace system by the Ringlace Systems, Inc., polyvinyl chloridene (PVCE) strings on racks are installed in the aeration tank. Each string has numerous loops of the same material, thus greatly increasing the surface area available for the growth of fixed-films (4). The Monitor process developed by the KLV Technologies is a system that consists of a media-filled biochamber, and wastewater is pumped into and through the aerated media. This design is suitable for upgrading lagoon installations for efficient nitrification (4). In addition, Kazuaki et al. (97) used hydrophobic porous membrane or oxygen enrichment membrane as a substratum of biofilm, i.e., biofilm formed on the oxygen permeable membrane. In this system, oxygen is diffused from the bottom to the surface of biofilm through the membrane, whereas organic pollutants are counter-diffused from the surface to the bottom of the biofilm. The membrane aeration allows nitrifying bacteria near the bottom region to grow with less competition with carbon oxidizers. As a result, simultaneous organics removal and nitrification can be achieved inside the biofilms. To enhance wastewater nitrification, enriched nitrifying bacteria are entrapped in a special biocarrier and further immobilized by sodium alginate to form spherical pellets with mean diameters of 1.0 to 2.0 mm (98). In addition, zeolite, a natural ion exchanger of ammonium ion, is coimmobilized into the pellets to enhance the efficiencies of transferring ammonium into the pellets in a batch fluidized-bed reactor (98). This modified nitrification technology seemed to be effective in treating wastewaters containing high-strength ammonia. Rostron et al. (99) compared PVA-encapsulated nitrification process with the Linpor system

under a variety of operating conditions. Results showed that the PVA-encapsulated nitrification process exhibited the highest nitrification rates under all testing conditions. This implies that PVA particles would be promising media, and are amenable to further optimization. A porous polyurethane carrier had also been applied for microbial immobilization to simultaneously remove carbonaceous and nitrogenous substances in fill-and-draw as well as in continuous-flow systems (100). In simultaneous organic removal and nitrification processes, inorganic carbon must be supplemented to stimulate the growth of autotrophic nitrifying bacteria. By adding sodium bicarbonate as an external inorganic carbon source, the oxidation of ammonium proceeded remarkably in the porous particle fluidized bed reactor (100).

A three-phase circulating floating bed reactor, namely TURBO N® had been developed by the company Degremont, and this reactor combined the advantages of high nitrification rates, no clogging, good mass transfer, and simple design and operation (101). Field study showed that TURBO N® could guarantee high nitrification rates and operation stability either in tertiary (up to 2 kg N/m^3 d) or secondary (up to 0.6 kg N/m^3 d) nitrification. This provides an attractive solution for intensive wastewater treatment for nitrogen and carbon removal. The prominent points of this reactor are such that (1) the separation of the reactor are completed in two sections, i.e., an up-flow aerated section and a down-flow nonaerated section, (2) a floating media with a high hold-up up to 40% apparent v/v is introduced, and (3) a homogeneous three-phase circulation (liquid-gas-solids) is induced by the injection of air. In fact, TURBO N® has the advantages of easier effluent and air-flow distribution, no primary settling, and no back-washing.

5.1.2. Separate-Stage Nitrification

Separate-stage nitrification may be suspended-growth-or attached-growth-based processes. The main concept of separate-stage nitrification is that carbonaceous oxidation and nitrification occurs in two or more separate biological units. Such a configuration allows for separately optimizing individual operation units, and has greater process flexibility and reliability. For example, each operation unit can be run independently to achieve optimum performance. On the other hand, in the separate-stage nitrification process, potential toxic effects may also be reduced because biodegradable organic materials, which may be toxic to nitrifying bacteria, are removed in the carbon oxidation stage.

The separate-stage suspended-growth nitrification processes are similar in design to the activated-sludge process. The degree of organic carbon removal in the carbon oxidation stage will affect the selection and operation of the nitrification unit. Low organic carbon in the influent may cause an imbalance between the solids lost from the sedimentation basins and the solids synthesized in the suspended growth reactor. Two different types of attached-growth processes, trickling filters and rotating biological contactors, have been used frequently for separate-stage nitrification. The packed-bed reactor has also been used, but only in a few applications.

5.2. Biological Denitrification Process

The conversion of nitrite and nitrate to nitrogen gas can be accomplished through a series of reduction reactions under anoxic conditions. Assimilatory and dissimilatory enzyme systems

are also involved in the reduction of nitrate. In the assimilatory nitrate reduction process, nitrate is converted to ammonia nitrogen for the use of cells biosynthesis; this occurs when nitrate is the only form of nitrogen available in culture. In the dissimilatory nitrate reduction process, nitrate is finally reduced to nitrogen, leading to the denitrification of wastewater. In most biological nitrification/denitrification systems, the wastewater to be denitrified must contain sufficient organic carbon, which is the energy source for the conversion of nitrate to nitrogen gas by bacteria. The organic carbon can be supplied through internal sources, such as wastewater and cell materials, or by the addition of an external source such as methanol. Denitrification can be accomplished in combined carbon oxidation nitrification/denitrification systems using internal and endogenous carbon sources or in separate reactors using methanol or other suitable external organics sources. The first process is called a single-stage sludge system, whereas nitrification/denitrification processes using separate reactors are often called separate or two-stage sludge systems.

5.2.1. Single-Stage Sludge Process

A single-stage sludge process for nitrogen removal basically combines carbonaceous removal, ammonia oxidation, and nitrate reduction within the same reactor. Different reactor configurations have been developed with various combinations of single or multiple anoxic zones, oxidation ditches, sequencing batch reactors, and cyclical aeration systems. In the single-stage sludge process, intermediate clarifiers or separate denitrification units are not necessary. As a result, there is a potential cost advantage as compared to the two-stage sludge system. However, it should be realized that the single-stage sludge process would have some potential technical limitations, such as high sensitivity to toxicity or inhibition owing to no separate upstream biological treatment step, lower nitrogen removal efficiency, difficulties in process control and optimization, relatively high energy usage, larger reactor volume and larger site requirements.

The most common configuration to achieve denitrification is to recycle nitrified mixed liquor to an antecedent anoxic zone, where exogenous carbon present in the influent wastewater can be used by the facultative denitrifying bacteria. Nitrates that are not recycled are discharged to the final clarifier. Anaerobic/anoxic/oxic (A^2/O), Modified Ludzake-Ettinger (MLE), Virginia Initiative Plant (VIP), and University of Capetown (UCT) processes belong to this category. It must be emphasized that process configurations using endogenous carbon for denitrification are not generally applied at full scale.

A proprietary single anoxic zone configuration is the A^2/O process (4), patented by the Air Products, Inc. Nitrification-denitrification is accommodated by the addition of an anoxic zone between the anaerobic and aerobic zones. Although the anaerobic zone is not required for nitrification-denitrification, it may be used at the start of the treatment train as an anaerobic selector to promote the proliferation of zoogleal organisms, and suppress the growth of filamentous bacteria in the anoxic and aerobic reactors. The MLE system is the improved Ludzack-Ettinger process, which places the anoxic denitrification zone ahead of the aerobic zone and uses external carbon in the raw wastewater. Meanwhile, an additional internal MLSS recycle from the aerobic stage to the anoxic stage is designed to return nitrified sludge at a regulated rate, which in turn ensures adequate nitrates for the heterotrophic denitrification

population. The interference of nitrates on phosphorus removal has been widely observed in the MLE and A^2/O processes. To solve this problem, the University of Capetown in South Africa developed the UCT process, in which activated sludge is returned to the anoxic zone instead of the anaerobic zone, and an additional recycle from the anoxic zone to the anaerobic zone was implemented (4). The purpose of these modifications is to denitrify the returned nitrates from the activated sludge line before they are recycled to the anaerobic zone.

The VIP process further refines the UCT process to accommodate lower strength wastewaters (4). Although the VIP and UCT processes seem to be similar, there are two fundamental differences. Firstly, the VIP process uses multiple complete mix cells instead of a single anaerobic unit to enhance phosphorus uptake by allowing a higher concentration of residual organics in the first anaerobic cell. Secondly, a shorter SRT is afforded in the VIP process to increase the proportion of active biomass in the mixed liquor. Multiple anoxic zones instead of a single unit have been developed for nitrification-denitrification purposes. For instance, the Barnard process uses a second anoxic zone for denitrification, whereas the modified UCT process has two anoxic zones (instead of one as in the original UCT) and two separate internal recycle lines. The purpose of the modified UCT is to control the return sludge and the nitrate recycle separately, and also to reduce the nitrate load to the anaerobic reactor.

Cyclical technologies are generally a modification of the activated sludge process. Alternating aerobic and anoxic zones can be achieved in a continuous-flow, activated sludge system by switching the aerators on and off (4). This type of intermittent or pulsed aeration in an activated sludge process is termed cyclical nitrogen removal (CNR), which can be effectively applied at plants with revised permits for nitrogen removal. An innovative alternating cyclical aeration process for nitrification-denitrification using countercurrent aeration is the Schreiber process. Alternating anoxic-aerobic zones within a single reactor is controlled by transferring air through submerged diffusers attached to a rotating arm (4). The mixed liquor typically rotates at a velocity less than the moving bridge. The moving diffuser concept is intended to prevent bubble rise in a common vertical path and to prevent inducement of vertical currents. Anoxic conditions can be achieved in the zone in front of the moving diffusers, while aerobic conditions exist in the zone immediately after the diffusers pass by that zone.

Oxidation ditches are perhaps the simplest treatment scheme for single sludge process. Anoxic conditions are achieved among the aerators as oxygen is depleted. The length of flow path per unit area can be increased by either arranging the flow loops concentrically or by folding the flow oval in half, namely the Orbal arrangement and the Carrousel process. The nitrification-denitrification option of the patented Orbal process by Envirex is termed the Sim-Pre process. The Sim-Pre process incorporates an internal recycle from the innermost to the outermost channel, which can serve as a predenitrification unit. In general, the simultaneous nitrification-denitrification phase occurs in the first aeration channel. Because the aeration demand exceeds the supply, anoxic conditions would be attainable along the flow path in the area upstream of the aerators. Thus, the outer channel of the Sim-Pre process is operated in a way similar to a conventional oxidation ditch.

The patented BioDenitro process employs multiple ditches for nitrogen removal, whereas the BioDenipho is a modified BioDenitro process for the simultaneous removal of phosphorus and nitrogen (14). Two configurations have been developed for the BioDenitro process. The

basic BioDenitro configuration consists of two identical aeration ditch tanks and a clarifier. Instead of creating anoxic and aerobic zones within each tank as in a conventional oxidation ditch, alternative aerobic or anoxic conditions are achieved within each looped reactor. Wastewater is fed alternatively between the two tanks to provide a carbon source for the desired microbial reactions. As compared to its basic process, the T-ditch configuration is designed in the second BioDenitro process with a six-phase cycle, in which first and third oxidation ditches are primarily used for denitrification and settling; the middle oxidation ditch serves as an aeration and flow distribution unit.

The sequencing batch reactor (SBR) is an important means for nitrification-denitrification and phosphorous removal. In SBR, denitrification occurs during the fill or react stages by alternatively switching aerators on and off, or during the settling and withdrawing periods. Several process innovations have been developed to enhance treatment. The Aqua SBR system developed by Aqua-Aerobic Systems, Inc., includes a proprietary floating direct drive mixer, an effluent decanter, and a microprocessor control system (4). The floating decanter is designed to prevent suspended solids from entering the decanter during mixed or react phases. In this SBR system, the supernatant below 30 cm of the water surface is withdrawn to mitigate scum losses to the effluent. The SBR designed by Omniflo-Jet Tech, Inc. is a proportional aeration system, in which the aeration capacity requirements are related to the volumetric change rate during the fill phase by sensing the DO level in the reactor (4). Such a strategy can help to optimize nitrification and denitrification cycles. The Cyclic Activated Sludge System (CASS) was developed by Transenviro, Inc., and is based on a conventional sequencing operation. This system is equipped with a captive selector, a baffled compartment, in which raw wastewater or primary effluent is mixed with returned sludge or internally recycled suspended solids. The mixture is then conveyed to the main reactor. Anoxic conditions can be created by limited or eliminated aeration in the selector, while concurrent high substrate levels are maintained. This mode of operation favors denitrification as well as the propagation of floc-formers.

Immobilized cell technology has been widely used in combined nitrification-denitrification process. To continuously remove the nitrogen from wastewater, a wide variety of carrier materials have been developed for use, and high nitrifying and denitrifying capacities can be achieved in the fixed-film reactors (6, 9, 102–105). A unique biological treatment system containing two membrane modules in a single tank was developed for simultaneous nitrification and denitrification (106). Both modules were fed with the substrates on the tube side of the silicone tubes by diffusing them to the biofilms formed on the tube surfaces. One module was fed with methanol for denitrification, and another was supplied with pure oxygen for nitrification. As the result, the interference of organic carbon on nitrification as well as the effect of dissolved oxygen on denitrification was hindered by the diffusion barriers (biofilms), thereby allowing nitrifying and denitrifying bacteria to coexist in a single tank.

Many existing treatment facilities have adopted a biological filter process for final effluent polishing because of its flexibility in combining with other treatment processes as well as for its compactness. For complete nitrogen removal, two separate submerged filters connected in series for nitrification and denitrification have been proposed (107). Nitrified effluent from the first aerated filter is mixed with the influent wastewater, and then fed to the second nonaerated

filter for further organics removal and denitrification. The two-filter system had been simplified by incorporating aerobic and anoxic/aerobic processes in a single filter (7, 107).

5.2.2. Separate-Stage Denitrification Process

In the separate-stage denitrification process, carbon oxidation, nitrification and denitrification are accomplished in separate reactors, and specific functional sludge is generated in each reactor. A supplemental carbon source is required in the separate-stage denitrification systems because almost all degradable organics present in the influent are removed in the carbon-oxidation and nitrification steps. The ideal supplemental organics should be inexpensive, readily available, essentially free of nitrogen and easily degradable. Neither raw sewage nor primary effluent is generally suitable as an external carbon source because of their high levels of ammonia, organic nitrogen, and suspended solid. For industrial scale treatment systems, methanol is the most appropriate choice and has been used because of its availability, low cost, favorable sludge production, low volatile organic compound emissions, and lack of nitrogen and phosphorus. In general, the separate-stage denitrification systems can also be subdivided into suspended growth systems and attached-growth systems. The design of separate stage suspended-growth denitrification system is similar in many respects to the design of the activated-sludge process for the removal of organic carbon. An aeration basin must be employed after the denitrification in anoxic reactor to strip out the nitrogen gas bubbles produced during the denitrification as well as to further oxidize methanol and organics remained.

Fixed-film processes have been used for separate-stage denitrification, such as the packed-bed reactor, circulating materials reactor, fluidized-bed reactor and RBC. In a fluidized-bed reactor, for instance, nitrified secondary effluent passes upward through a column at a flow rate sufficient to produce a fluidized bed of media (typically sand) on which denitrifying bacteria attach. Because the media are fluidized and the particles are not in contact with each other, extremely large specific surface area can be achieved for the attachment of growth of denitrifying bacteria. The specific surface area available for biological growth is about 10 times higher than the downflow packed-bed system.

6. COMMERCIALIZED NITROGEN REMOVAL PROCESSES

To date, several innovative nitrogen removal processes have been commercialized, most of them biofilm-based systems. DeepBed™ sand filter is a fixed-film biological process for organics oxidation, nitrification and denitrification, and microorganisms are encouraged to grow on the surface of sand media (108). In the aerobic CoIOX™ unit, wastewater and air flow upward through the media bed where the microorganisms uptake oxygen from air to oxidize organics and ammonium, whereas in the anoxic Denite® process, nitrified wastewater fed with a carbon source flows downward through the filter bed. In practice, biofilters can play a double role of bioreactor and filter simultaneously without the need of an additional clarifier. Therefore, biological treatment, clarification, and filtration of wastewater can be accomplished in a single reactor.

A suspended carrier technology is currently used to enhance nitrification and denitrification within the existing aeration tank. Suspended carrier or moving bed processes for attached

growth are a recent innovation in the field of domestic and industrial wastewater treatment for removal of organic matter as well as nutrients. As a bacterial biofilm forms on the carriers that are suspended in the aeration tanks, the total biomass concentration in the aeration tanks can be increased, whereas the increased biomass concentration, and particularly the much higher SRT of the attached biomass, results in efficient carbon removal and improved nitrification and dentrification through process intensification. The media commonly used are freely floating carriers that can be made of different materials, shapes and sizes (9, 109). Because almost the entire biomass is located on the carriers and only a very small fraction grows in suspension, the resulting suspended solid concentration of the effluent is low. Subsequently, the clarifier volume can be kept to a minimum. In Europe, the first full-scale ASTRASAND moving bed biofiltration process for biological postdenitrification was reported by Kramer et al. (110).

One solution for upgrading the conventional activated sludge treatment plant to remove nitrogen is to implement a biofilter system downstream of the existing plant (111). A typical example is the BIOSTYR system, which is an upflow biofilter with a flowing filter bed of expanded polystyrene beads. The activated sludge plants upgraded with the BIOSTYR system have successfully operated for tertiary nitrification in France. An airlift reactor has been used for the removal of organics and nutrients. An airlift-based CIRCOX® system had been scaled up and used for nutrient removal (112). In this patented reactor, the sludge on carrier is circulated between the oxic and the anoxic compartment by means of an airlift pump.

The Mixazur® is a patented fixed-film mobile bed reactor for denitrification, which combines the advantages of fluidized-bed, fixed-bed reactors, and activated sludge process (113). In the Mixazur® system, the first zone is a reaction zone in which optimized contact is created among the bioparticles, a mixture of wastewater and returned nitrified liquor, whereas the mixture of bioparticles and wastewater is maintained in suspension by mechanical agitation. A relatively high axial speed results in a sufficient shear stress leading to an efficient control of a thin and active biofilm. The second zone in the system is the separation zone, in which the sedimentation of bioparticles from the treated effluent is achieved. The excess free biomass detached from the bioparticles is continuously carried away with the effluent towards the nitrification stage.

The B2A biofilter is designed in vivo as a predenitrification biofilter for complete nitrogen removal (113). The process consists of an upflow filtration of prescreened raw wastewater through a series of decreasing sized granular media of 80 to 2.5 mm. Because the prescreened wastewater is fed to the anoxic part, a large amount of particulate organic matter can be accumulated in the lower part of the filter between each backwash. When the head loss is above a certain level, the filters are backwashed in two steps: (1) a gravitational purge is used to backwash the lower layer of the filter where the biggest particles are situated; (2) air and water are injected at a high pressure in the bottom of the filter to remove biomass and particulates from the middle and upper part of the filter. During the backwash, the filter materials in the middle and upper part are elevated, but not mixed completely. Hence, no significant scouring of the filter media occurs. The anoxic and the aerobic parts of the B2A biofilter are situated in the same column with the anoxic part below (114).

7. NEW BIOLOGY FOR NITROGEN REMOVAL

Traditional biological nitrogen removal processes have high oxygen and energy demands. Recently developed processes can make nitrogen removal more sustainable and reduce organics requirements and energy consumption.

7.1. Nitrite Route

Direct denitrification from the nitrite stage by performing a nitrate shunt can achieve advantages such as a reduction of carbon requirements, lower energy consumption for aeration, reduced reactor volume owing to a shortened reaction pathway, as well as significant reduction in plant operation costs. To assess the feasibility of a shortened pathway for nitrogen removal, it is necessary to have a better understanding of the nitrification kinetics and then to find a way to promote nitrite build-up in a biological nitrification system (6, 46, 50). At least three factors have been found to influence nitrite build-up: (1) the relative specific growth rates of *Nitrosomonas* to *Nitrobacter*, μ_{Ns}/μ_{Nb} in the biofilm; (2) the relative initial ratio between *Nitrosomonas* and *Nitrobacter* on the support surface, $(M_{ao})_{Ns}/(M_{ao})_{Nb}$; (3) the level of free ammonia, particularly at greater than 0.1 mg N/l that can be inhibitory to *Nitrobacter* (46). In addition, the washout of nitrite oxidizers based on growth rate is another possible control option for nitrite build-up (7, 46). At elevated temperatures ($>15°C$), the ammonium oxidizers have a much higher growth rate than the nitrite oxidizers do. Hence, carefully controlling the sludge age has been shown to be a good operating strategy for a stable partial nitrification (50, 115). In addition, the affinity of nitrite oxidizers to oxygen is much lower than ammonium oxidizers; this can be used to selectively restrict the growth of nitrite oxidizers by manipulating the concentration of dissolved concentration (116).

SHARON is the process with single reactor system for high-rate ammonium removal over nitrite (117). In this process, nitrification is significantly enhanced because ammonium oxidizers have a higher relative growth rate than nitrite oxidizers under proper conditions. Essentially, the SHARON is a chemostat system, in which the dilution rate is controlled at level higher than the maximum growth rate of nitrite oxidizing bacteria, but lower than the growth rate of ammonium-oxidizing bacteria. Such an operation strategy favors the accumulation of nitrite or partial nitrification occurring in the reactors. The SHARON process has been successfully operated in combination with denitrification to treat high-strength nitrogen-containing wastewater.

7.2. Aerobic Denitrification

Studies by Patureau et al. (118) show that oxygen and nitrate can be consumed simultaneously by some microbial strains in a phenomenon called corespiration. These properties of corespiration can allow nitrifiers and aerobic denitrifier (e.g., *Microvirgula aerodenitrificans*) to coexist in a single aerated reactor under a continuous or sequencing batch reactor. The aerobic denitrifier can be maintained by the intermittent addition of organic carbon.

7.3. Autotrophic Denitrification

Denitrification by nitrifiers had been proposed by Wrage et al. (119), namely nitrifier denitrification. In this pathway of nitrification, ammonia (NH_3) is oxidized to nitrite (NO_2^-) followed by the reduction of NO_2^- to nitric oxide (NO), nitrous oxide (N_2O) and molecular nitrogen (N_2). These transformations are carried out only by autotrophic nitrifiers. Thus, it seems that nitrifier denitrification differs from the coupled nitrification-denitrification, in which denitrifiers are responsible for the reduction of NO_2^- or NO_3^- produced by nitrifiers to nitrogen gas. In fact, nitrifier denitrification mainly occurs in soils, and low oxygen conditions coupled with low organic carbon contents of soils favor this pathway. Finally, it should be pointed out that nitrifier denitrification would contribute to the production of the greenhouse gas N_2O, and also causes losses of fertilizer nitrogen in agricultural soils.

Besides nitrifier denitrification, other autotrophs can also cause denitrification. An innovative process using sulfur-limestone autotrophic denitrification, SLAD for short, was developed for the treatment of nitrate-contaminated surface or wastewater under both aerobic and anaerobic conditions (120). Both autotrophic denitrifiers and nondenitrifying bacteria, such as *Thiobacillus thiooxidans* were detected in the process under aerobic conditions. Because no organic carbon source is needed in the SLAD process, and autotrophic denitrifiers exist widely in natural sediments or soil, the SLAD process may be considered as a replacement for heterotrophic denitrification in constructed wetlands or stabilization ponds. The SLAD process has been extended by incorporating a membrane separation unit. Autotrophic denitrifiers, whose growth rates are considerably low, can be kept at a high concentration by the membrane separation. A rotating membrane disk module equipped with a UF membrane was incorporated in the SLAD system (121).

7.4. Heterotrophic Nitrification

In addition to the autotrophic nitrification, many heterotrophic bacteria are able to produce oxidized nitrogen forms from ammonia. After heterotrophic nitrification was suggested in 1894, many heterotrophic bacteria such as *Arthrobacter globiformis, Aerobacter aerogenes, Mycobacerium phlei, Stretomyces griseus, Thiosphaera*, and *Pseudomonas* spp. have been found to have nitrifying functions (50). In contrast to the autotrophic nitrification that is proportionally related to cell growth, heterotrophic nitrification is indeed independent of cell yield. This is owing to the fact that most of the products of heterotrophic nitrification are formed during the stationary growth phase, and the heterotrophic nitrification reactions are not ATP-coupled. Castignetti (122) studied the proton translocation of heterotrophic nitrifiers, and found that heterotrophic nitrification did not conserve energy during the oxidation of nitrogenous substrates. Although heterotrophic nitrification has been demonstrated in soils, sewage treatment, rivers and lake waters, autotrophic nitrification is over ten times more significant than heterotrophic nitrification in natural systems (Table 13.2) (79). When substances that are selectively inhibitory to autotrophic nitrifiers are added to soil or activated sludge, nitrification is usually completely inhibited. These indicate that autotrophic nitrification is the main oxidation pathway of ammonium, and heterotrophic nitrification does not seem to make a major contribution to the conversion of ammonia to nitrite and nitrate ions.

Table 13.2
Rates of nitrification by some heterotrophic and autotropic nitrifiers

Organism	Substrate	Product	Max. product accumulation (μg N/mL)	Rate of formation (μg N/g dry cells d)
Aspergillus (heterotroph)	NH_4^+	NO_2^-	75	1349
Arthrobacter (heterotroph)	NH_4^+	NO_2^-	0.2–1.2	370–9000
Arthrobacter (heterotroph)	NH_4^+	NO_2^-	2–4.4	250–655
Nitrosomonas (autotroph)	NH_4^+	NO_2^-	2000–4000	1–30 million
Nitrobacter (autotroph)	NO_2^-	NO_2^-	2000–4000	5–70 million

Adapted from (79).

In addition to the heterotrophic nitrification discussed earlier, some heterotrophic bacteria, such as *Thiosphaera pantotropha*, can contribute to simultaneous nitrification and denitrification (123, 124). *Thiosphaera pantotropha* can oxidize ammonia heterotrophically to nitrite and nitrate, and further reduce them to nitrogen gas irrespective of the ambient dissolved oxygen concentration. To date, *Thiosphaera pantotropha* has been tested in suspended and attached mixed cultures for the treatment of industrial and domestic wastewaters (125). A three-stage rotating biological contactor (RBC) was developed with a mixed culture of *Thiosphaera pantotropha*, autotrophic nitrifiers and other heterotrophs (126). Contrary to the conventional RBC units designed for a concurrent carbon removal and nitrification, the nitrification rate in the RBC system associated with heterotrophic nitrification-denitrification increased linearly with an increase in organic loading rate before stabilizing at a COD loading rate of about $15\,kg/m^2 \cdot d$ and $1.5\,kg\,N/m^2 \cdot d$ for nitrogen. These seem to indicate that a single-stage aerobic biofilm reactor augmented with *Thiosphaera pantotropha* could meet the increasingly stringent regulations on effluent nitrogen discharges, and also affords several advantages over the conventional systems: low buffer requirements, no need for external carbon source for denitrification and so on. Consequently, a substantial reduction in the treatment cost would be expected.

7.5. Anaerobic Ammonium Oxidation (Anammox)

The anaerobic oxidation of ammonium by deep-branching *Planctomycetes*, which can be used for cost-effective and space-saving nitrogen removal from high-strength wastewater, had been widely reported in the environmental engineering field (127–129). In 1986, a special reaction zone called Anammox (anaerobic ammonium oxidation) was discovered in an anaerobic denitrifying fluidized bed reactor (130). During the anaerobic ammonium oxidation, ammonium and nitrite are directly converted to dinitrogen gas without requiring COD or the addition of an external carbon source as follows (131):

$$NH_4^+ + NO_2^- \rightarrow N_2 + 2H_2O.$$

In this reaction, nitrite is the preferred electron acceptor, whereas hydroxylamine and hydrazine are identified as important intermediates. Under suboptimal conditions, the overall double time of microbial community responsible for the anaerobic ammonium oxidation is longer than 3 weeks (128). The Anammox activity is highly sensitive to oxygen, and even

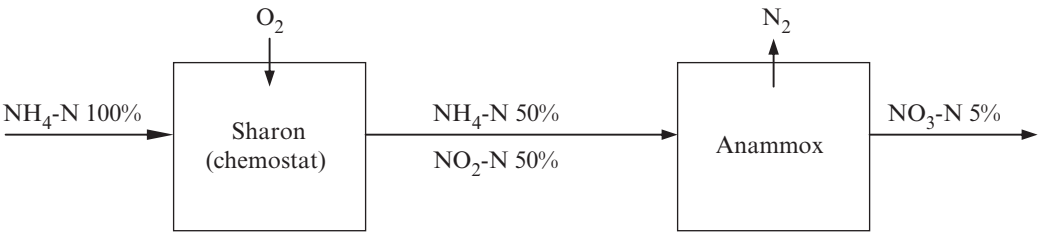


Fig. 13.6. Schematic representation of the combined Sharon-Anammox process for the removal of ammonium from sludge digestion effluents (adapted from (134)).

a trace amount of oxygen of 0.03 mg/L could inhibit this process (132). However, once the oxygen is removed or depleted, the anaerobic ammonium oxidizing bacteria can resume their metabolism. This obligated anaerobic nature is in sharp contrast to the versatile metabolism of aerobic ammonium-oxidizing bacteria (133). Anammox can also be inhibited by nitrite but not ammonium and nitrate, i.e., the Anammox process may be inactivated at a nitrite concentration greater than 100 mg N/L (134).

The combination of the Anammox process and a partial nitrification process, SHARON for example, allows ammonium to directly be converted to dinitrogen gas. It is expected that in such a combined system, the problems encountered in traditional denitrification process (needs of an extra electron donor) can be circumvented (135). In the combined Anammox and SHARON process (Figure 13.6), equal amounts of ammonium and nitrite are produced in the SHARON reactor, then this mixture of ammonium and nitrite is fed into an Anammox SBR. Compared to conventional systems, the combined Anammox and SHARON process has advantages of low sludge production less energy and oxygen inputs requirements, and no need for external carbon addition. This combination can be realized in biofilm reactors (132, 136). It can be expected that this combined process would be a promising biotechnology for handling wastewater with a high ammonium and low BOD content in near future. Qualitative comparison of several components of the Anammox technology with conventional nitrogen removal system is shown in Table 13.3.

7.6. New Metabolisms

Cometabolism is a widespread process in nature. The cometabolic removal of recalcitrant organics by nitrifiers has been used in the wastewater treatment process. Nitrifying activity may be associated with the generation of OH^- radicals by the ammonia monooxygenase (AMO) (137), and is related to the production of soluble microbial products that can act as cosubstrates for heterotrophic bacteria. The action of the nitrifiers, which can be stimulated by supplying a specific substrate, i.e, NH_3, can in this way bring about the initiation of indirect biodegradation of recalcitrant organics. The enhancement of ethane removal in a packed granular activated carbon biobed was achieved by specifically stimulating nitrification (138).

The use of molecular techniques can construct more efficient degradative strains. The in vivo assembly of partial catabolic sequences from different pathways and different organisms is a practical strategy to the evolution of new complete catabolic pathways. Genetic

Table 13.3
Qualitative comparison of several components of the anammox technology with conventional nitrogen removal system

System	Conventional nitrification-denitrification	Sharon	Anammox	CANON
Number of reactor feed	2	1	1	1
Reactor capacity (kg N/m^3 day)	0.05–4	1	6–12	1–3
Discharge	NO_3^-; N_2O, N_2	NH_4^+, NO_2^-	N_2, NO_3^-	N_2, NO_3^-
Conditions	oxic; anoxic	oxic	anoxic	oxygen limited
pH control	yes	none	none	none
Biomass retention	none	none	yes	yes
COD requirement	yes	none	none	none
Oxygen requirement	high	low	none	low
Sludge production	high	low	low	low
Bacteria	Nitrifiers + various heterotrophs	Aerobic NH_4^+ oxidizers	Planctomycetes	Aerobic NH_4^+ oxidizers + planctomycetes

Adapted from (131).

engineering can be a powerful technique to help to establish effective degradation pathways, and further to accelerate the evolution of totally new degradative capabilities (139). This allows the precise selected genes that are essential for the hybrid pathway to become possible. Although no report has been presented for the use of new genetically constructed pathways in nitrification and denitrification process, advanced technologies will be soon appear with the development of molecular techniques and genetic engineering.

8. NEW FINDINGS OF BACTERIA FOR NITROGEN REMOVAL

In the past few years, the application of advanced molecular techniques, such as FISH, competitive PCR, and 16S rRNA gene analysis, has greatly enriched the understanding of bacteria responsible for nitrogen removal. Previously unrecognized bacteria were demonstrated to catalyze nitrogen removal in activated sludge and biofilm processes (11, 25). To date, many text or professional technical books still report that *Nitrosomonas* and *Nitrobacter* are the main ammonium and nitrite oxidizers in wastewater treatment plants (60, 140). However, by using several the modern molecular methods, it becomes evident that previously uncultured *Nitrospira*-like bacteria rather than *Nitrobacter* spp., are the dominating nitrite oxidizers in most wastewater treatment plants across the world (141, 142). In a submerged membrane aerobic reactor, Witzig et al. (25) found that neither *Nitrosomonas* nor *Nitrosospira*, were present in the membrane filtration sludge. On the other hand, there is evidence to show that nitrite oxidation in activated sludge would not be related to *Nitrobacter* spp. (20). Consequently, *Nitrospira* spp. might play a more important role in nitrite oxidation than the genus *Nitrobacter*. These molecular techniques-based findings, without any doubt, would profoundly change the actual picture of biological nitrification-denitrification, and lead to the rapid development of innovative biotechnology for efficient removal of nitrogen from wastewater (143, 144).

9. DESIGN EXAMPLE

Design a complete-mix single stage activated sludge process to treat $20{,}000\,m^3/d$ primary effluent to meet biodegradable COD concentration of $1.0\,g/m^3$, NH_4-N concentration of $0.40\,g/m^3$ in the final effluent from the secondary clarifier. The temperature is 20°C. Following information is available for the design.

Wastewater characteristics

Constituent	Concentration	Unit
Biodegradable COD, $(COD)_b$	230	g/m^3
BOD	144	g/m^3
TSS	63	g/m^3
VSS	53	g/m^3
Nonbiodegradable *VSS*, $(VSS)_{nb}$	20	g/m^3
NH_4-N	25	g/m^3
Flowrate, Q	20,000	m^3/d
Alkalinity	129	g/m^3 as $CaCO_3$
$(COD)_b$/BOD ratio	1.6	

Kinetic coefficients for nitrifying bacteria

Coefficients	Value	Unit
Maximum specific growth rate, $\mu_{n,m}$	0.50	d^{-1}
The NH_4-N-based Monod constant, K_n	0.60	g/m^3
Endogenous decay rate, k_{dn}	0.05	d^{-1}
Biomass yield, Y_n	0.10	g VSS/g NH_4-N

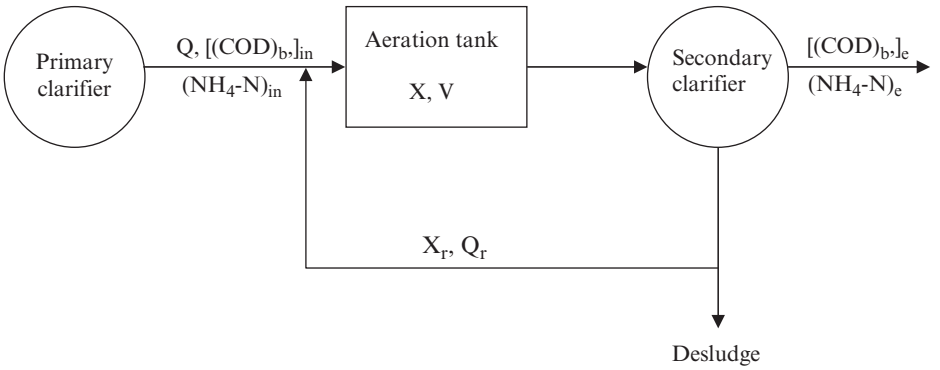
Kinetic coefficients for heterotrophic bacteria

Coefficients	Value	Unit
Maximum specific growth rate, μ_m	3.5	d^{-1}
Endogenous decay rate, k_d	0.088	d^{-1}
The organic-based Monod constant, K_c	18	g/m^3
The oxygen-based Monod constant, K_o	0.60	g/m^3
Biomass yield, Y	0.35	g VSS/g $(COD)_b$

Design assumptions:

(a) *DO* in aeration basin $= 2.0\,g/m^3$
(b) Safety factor $= 1.2$
(c) Fraction of biomass that remains as cell debris, $f_d = 0.15$
(d) Design total suspended solid (*TSS*) concentration, $X_{TSS} = 2800\,g/m^3$
(e) Soluble BOD and *TSS* in the effluent are $3.0\,g/m^3$ and $8.0\,g/m^3$ respectively

Notes: This example is analogue to that given by Tchobanoglous et al. (143). Calculations presented here also follow the step-by-step design guidelines proposed by Tchobanoglous et al. (143).



(1) Determine the specific growth rate μ_n for nitrifying bacteria

$$\mu_n = \left(\frac{\mu_{n,max}(NH_4\text{-}N)_e}{K_n + (NH_4\text{-}N)_e}\right)\left(\frac{DO}{K_o + DO}\right) - k_{dn}$$

$$= \left(\frac{(0.50\,g/g{\cdot}d)(0.40\,g/m^3)}{(0.60\,g/m^3) + (0.40\,g/m^3)}\right)\left(\frac{2.0\,g/m^3}{(0.60\,g/m^3) + (2.0\,g/m^3)}\right) - (0.05\,g/g{\cdot}d)$$

$$= 0.10\,g/g{\cdot}d$$

(2) Determine the theoretical *SRT* for nitrifying bacteria

$$SRT = \frac{1}{\mu_n} = \frac{1}{0.10\,\text{g/g·d}} = 10\ \text{days}$$

Given the safety factor of 1.2, then design *SRT* = safety factor × theoretical *SRT* = 1.2 × (10 d) = 12 d

(3) Determine biomass synthesized (P_{bio})

$$P_{bio} = \frac{QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}}{1 + k_d \times SRT} + \frac{(f_d)(k_d)QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}SRT}{1 + k_d \times SRT}$$
$$+ \frac{QY_n(\text{NH}_4\text{-N})_{in}}{1 + k_{dn} \times SRT}$$
$$= \frac{(20000\,\text{m}^3/\text{d})[0.35\,\text{gVSS/g(COD)}_b][(230 - 1.0)\,\text{g (COD)}_b/\text{m}^3]}{1 + (0.088\,\text{d}^{-1})(12\,\text{d})}$$
$$+ \frac{(0.15)(0.088\,\text{d}^{-1})(20000\,\text{m}^3/\text{d})[0.35\,\text{gVSS/g (COD)}_b][(230 - 1.0)\text{g(COD)}_b/\text{m}^3](12\,\text{d})}{1 + (0.088\,\text{d}^{-1})(12\,\text{d})}$$
$$+ \frac{(20000\,\text{m}^3/\text{d})(0.10\,\text{g VSS/g})(25\,\text{g/m}^3)}{1 + (0.05\,\text{d}^{-1})(12\,\text{d})}$$
$$= 934.4\,\text{kg VSS/d}$$

(4) Determine the amount of nitrogen oxidized to nitrate (NOx-N)

$$\text{NO}_x\text{-N} = (\text{NH}_4\text{-N})_{in} - (\text{NH}_4\text{-N})_e - 0.12P_{bio}/Q$$
$$= 25\,\text{g/m}^3 - 0.5\,\text{g/m}^3 - 0.12(934.4\,\text{kgVSS/d})(1000\,\text{g/kg})/(20000\,\text{m}^3/\text{d})$$
$$= 18.9\,\text{g/m}^3$$

(5) Determine the concentration and mass of *VSS* and *TSS* in the aeration basin

a) Calculate the concentration of *VSS* (P_{VSS}) and *TSS* (P_{TSS}) in the aerobic basin

$$P_{VSS} = \frac{QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}}{1 + (k_d)SRT} + \frac{(f_d)(k_d)QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}SRT}{1 + (k_d)SRT}$$
$$+ \frac{QY_n(\text{NH}_4\text{-N})_{in}}{1 + (k_{dn})SRT} + Q(VSS)_{nb}$$
$$= P_{bio} + Q(VSS)_{nb}$$
$$= 934.4\,\text{kg/d} + (20000\,\text{m}^3/\text{d})(20\,\text{g/m}^3)(1\,\text{kg}/1000\,\text{g})$$
$$= 1334.4\,\text{kg/d}$$
$$P_{TSS} = \left\{\frac{QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}}{1 + (k_d)SRT} + \frac{(f_d)(k_d)QY\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\}SRT}{1 + (k_d)SRT}\right.$$
$$\left. + \frac{QY_n(\text{NH}_4 - \text{N})_{in}}{1 + (k_{dn})SRT}\right\} \Big/ 0.85 + Q(VSS)_{nb} + Q(TSS_{in} - VSS_{in})$$
$$= P_{bio}/0.85 + Q(VSS)_{nb} + Q(TSS_{in} - VSS_{in})$$

$$= \frac{934.4\,\text{kg/d}}{0.85} + (20000\,\text{m}^3/\text{d})(20\,\text{g/m}^3)(1\,\text{kg}/1000\,\text{g})$$
$$+ (20000\,\text{m}^3/\text{d})(63-53)\,\text{g/m}^3(1\,\text{kg}/1000\,\text{g})$$
$$= 1699.3\,\text{kg/d}$$

b) Calculate the mass of *VSS* and *TSS* in the aerobic basin

$$\text{Mass of } VSS = (P_{\text{VSS}})SRT = (1334.4\,\text{kg/d})(12\,\text{d}) = 16{,}012.8\,\text{kg}$$
$$\text{Mass of } TSS = (P_{\text{TSS}})SRT = (1699.3\,\text{kg/d})(12\,\text{d}) = 20{,}391.6\,\text{kg}$$

Because *VSS* is much more than *TSS*, the mass of *TSS* is used to calculate the aeration tank volume.

(6) Determine the aeration tank volume (V)
Given a TSS concentration of $2{,}800\,\text{g/m}^3$, then $(X_{\text{TSS}})(V) = 20391.6\,\text{kg}$, that is,

$$V = \frac{(20391.6\,\text{kg})(1000\,\text{g/kg})}{2800\,\text{g/m}^3} = 7283\,\text{m}^3$$

Thus, use 2 aeration tanks with the volume of $3642\,\text{m}^3$ each.

(7) Determine the hydraulic retention time (*HRT*) in the aeration tank

$$HRT = \frac{V}{Q} = \frac{(7283\,\text{m}^3)(24\,\text{h/d})}{20000\,\text{m}^3/\text{d}} = 8.8\,\text{h}$$

(8) Determine *VSS* (X_{VSS})

$$\text{Fraction } VSS = \frac{16012.8\,\text{kg}}{20391.6\,\text{kg}} = 0.78, \text{ then } \text{X}_{\text{VSS}} = 0.78(2800\,\text{g/m}^3) = 2184\,\text{g/m}^3$$

(9) Determine *F/M* ratio and *BOD* volumetric loading rate
a) Determine *F/M* ratio

$$F/M = \frac{Q(\text{BOD})_{\text{in}}}{X_{\text{VSS}}V} = \frac{(20000\,\text{m}^3/\text{d})(144\,\text{g BOD/m}^3)}{(2184\,\text{g VSS/m}^3)(7283\,\text{m}^3)} = 0.18\,\text{g BOD/g VSS}\cdot\text{d}$$

b) Determine volumetric BOD loading rate

$$L_{\text{BOD}} = \frac{Q(\text{BOD})_{\text{in}}}{V} = \frac{(20000\,\text{m}^3/\text{d})(144\,\text{g BOD/m}^3)}{(7283\,\text{m}^3)(1000\,\text{g/kg})} = 0.40\,\text{kg BOD/m}^3\cdot\text{d}$$

(10) Determine the observed growth yield based on *TSS* and *VSS* (Y_{TSS} and Y_{VSS})

$$(\text{COD})_{\text{b}}\text{ removed} = Q\{[(\text{COD})_{\text{b}}]_{\text{in}} - [(\text{COD})_{\text{b}}]_{\text{e}}\}$$
$$= (20{,}000\,\text{m}^3/\text{d})(230-1)\text{g/m}^3(1\,\text{kg}/1000\,\text{g}) = 4580\,\text{kg/d}$$

Given $\text{BOD} = (\text{COD})_{\text{b}}/1.6$

$$\text{BOD removed} = \frac{4580\,\text{kg}(\text{COD})_{\text{b}}/\text{d}}{1.6\,\text{kg}(\text{COD})_{\text{b}}/\text{kg BOD}} = 2862.5\,\text{kg BOD/d}$$

Hence, the observed growth yield based on *TSS* can be calculated as follows:

$$Y_{\text{TSS}} = \frac{1699.3\,\text{kg TSS/d}}{2862.5\,\text{kg BOD/d}} = 0.59\,\text{kg TSS/kg BOD} = 0.59\,\text{g TSS/g BOD}$$

and the observed growth yield based on *VSS*

$$Y_{\text{VSS}} = \frac{1334.4\,\text{kg VSS/d}}{2862.5\,\text{kg BOD/d}} = 0.47\,\text{kg VSS/kg BOD} = 0.47\,\text{kg VSS/g BOD}$$

(11) Calculate the oxygen demand (R_o)

$$\begin{aligned} R_o &= Q\{[(\text{COD})_b]_{in} - [(\text{COD})_b]_e\} - 1.42 + 4.33\,Q(\text{NO}_x\text{-N}) \\ &= (20{,}000\,\text{m}^3/\text{d})(230-1)\text{g/m}^3(1\,\text{kg}/1000\,\text{g}) - 1.42(934.4\,\text{kg VSS/d}) \\ &\quad + 4.33(20000\,\text{m}^3/\text{d})(18.9\,\text{g/m}^3)(1\,\text{kg}/1000\,\text{g}) \\ &= 4889.9\,\text{kg/d} = 203.7\,\text{kg/h} \end{aligned}$$

(12) Check alkalinity
Alkalinity to maintain pH7 = Influent Alkalinity − Alkalinity used + Alkalinity to be added
Alkalinity needed to maintain pH in the range of 6.8–7.0 is about 70–80 g/m^3 as $CaCO_3$, thus a middle value of 75 g/m^3 $CaCO_3$ is selected for this design.

Given the influent Alkalinity: 129 g/m^3 as $CaCO_3$, and the amount of nitrogen converted to nitrate is 18.9 g/m^3, then alkalinity for nitrification = (7.14 g $CaCO_3$/g NH_4-N)(18.9 g NH_4-N/m^3) = 134.9 g/m^3 as $CaCO_3$. Hence, alkalinity required = (75 + 134.9 − 129)g/m^3 = 80.9 g/m^3 as $CaCO_3$ Alkalinity required daily = (80.9 g/m^3)(20,000 m^3/d)(1 kg/1000 g) = 1618 kg/d as $CaCO_3$

(13) Estimate the total BOD in effluent

$$\begin{aligned} \text{Total BOD} &= (\text{BOD})_{se} + \text{BOD owing to the presence of } \textit{VSS} \\ &= (\text{BOD})_{se} + \left(\frac{1\,\text{gBOD}}{1.42\,\text{g VSS}}\right)\left(\frac{0.85\,\text{g VSS}}{\text{g TSS}}\right)(\text{TSS, g/m}^3) \end{aligned}$$

Given $(\text{BOD})_{se}$ of 3.0 g/m^3, and *TSS* of 8.0 g/m^3, hence total BOD in the effluent = 3.0 g/m^3 + (0.70)(0.85)(8.0 g/m^3) = 7.8 g/m^3

(14) Design the secondary clarifier
Activated sludge return ratio (R):

$$R = \frac{X_{\text{TSS}}}{X_{r,\text{TSS}} - X_{\text{TSS}}}$$

Assume $X_{r,\text{TSS}} = 7500\,\text{g/m}^3$, then

$$R = \frac{2800\,\text{g/m}^3}{(7500-2800)\,\text{g/m}^3} = 0.60$$

By assuming a hydraulic application rate of 25 $\text{m}^3/\text{m}^2 \cdot \text{d}$ at average flow for the secondary clarifier, the surface area of clarifier can be computed.

$$\text{Surface area} = \frac{20000\,\text{m}^3/\text{d}}{25\,\text{m}^3/\text{m}^2 \cdot \text{d}} = 800\,\text{m}^2$$

Use 2 clarifiers with a diameter of about 22 m

(15) Check solids loading to the secondary clarifier

$$\text{Solids loading} = \frac{(Q + Q_r)(X_{TSS})}{A} = \frac{(1 + R)Q(X_{TSS})}{A}$$

$$= \frac{(1 + 0.6)(20000\,m^3/d)(2800\,g/m^3)(1\,kg/1000\,g)}{(800\,m^2)(24\,h/d)}$$

$$= 4.7\,kg\ TSS/m^2\,h$$

This value is within acceptable range of solids loading of 4 to $6\,kg/m^2 \cdot h$.

Summary of the design parameters

Parameter	Value	Unit
Average wastewater flow	20000	m^3/d
Aerobic *SRT*	12	d
Aeration tanks	2	number
Total aeration tank volume	7283	m^3
HRT	8.8	h
TSS	2800	g/m^3
VSS	2184	g/m^3
F/M	0.18	g BOD/g VSS · d
BOD loading	0.40	g $BOD/m^3 \cdot d$
Sludge production	1699.3	kg TSS/d
Overall growth yield	0.59	kg TSS/kg $(COD)_b$
	0.47	kg VSS/kg $(COD)_b$
Oxygen required	203.7	kg/h
Sludge return ratio	0.6	
Clarifier hydraulic application rate	25	$m^3/m^2 \cdot d$
Clarifiers	2	number
Diameter of each clarifier	22	m
Alkalinity addition as $CaCO_3$	1618	kg/d
Effluent total BOD	7.8	g/m^3
Effluent *TSS*	8	g/m^3
Effluent NH_4-N	0.4	g/m^3

NOMENCLATURE

μ_D = specific denitrifier growth rate, d^{-1}
$\mu_{D,max}$ = maximum specific growth rate of denitrifying bacteria, d^{-1}
μ_n = specific growth rate for the nitrifying bacteria, d^{-1}
$\mu_{n,max}$ = maximum specific growth rate of nitrifying bacteria, d^{-1}
μ_N = specific growth rate of ammonium oxidizer, d^{-1}
$\mu_{N,max}$ = maximum specific growth rate of ammonium oxidizer, d^{-1}
μ_m = maximum specific growth rate for heterotrophic bacteria, d^{-1}

μ_n' = net specific growth rate of nitrifying bacteria, d^{-1}
μ_D' = net specific growth rate of denitrifying bacteria, d^{-1}
Area = area of clarifier, m^2
BOD = biochemical oxygen demand, g/m^3
BOD_{se} = soluble BOD and TSS in the effluent, g/m^3
$(BOD)_{in}$ = biological oxygen demand in the influent, g/m^3
COD = chemical oxygen demand, g/m^3
$(COD)_b$ = biodegradable COD, g/m^3
$[(COD)_b]_{in}$ = concentration of biodegradable COD in the influent, g/m^3
$[(COD)_b]_{out}$ = concentration of biodegradable COD in the effluent, g/m^3
DO = dissolved oxygen concentration, g/m^3
F/M = food to biomass ratio, g BOD/gVSS · d
FA = free ammonia concentration, g/m^3
f_d = fraction of biomass that remains as cell debris
HRT = hydraulic retention time, hour
k_{dn} = decay rate of nitrifying bacteria, d^{-1}
k_d = endogenous decay rate heterotrophic bacteria, d^{-1}
K_D = half-saturation constant for denitrifying bacteria, g/m^3
K_N = half-saturation constant for ammonium oxidizer, g/m^3
K_n = NH_4-N based Monod constant, g/m^3
K_{NO3} = half-saturation constant for nitrate nitrogen, g/m^3
K_o = half-saturation constant for oxygen, g/m^3
K_s = half-saturation or half-velocity constant (equivalent to the growth-limiting substrate concentration at half the maximum specific growth rate), g/m^3
K_c = organic based Monod constant, g/m^3
L_{BOD} = volumetric BOD loading rate, kg BOD/m^3 · d
NH_4-N = NH_4^+-N concentration, g/m^3
NO_2-N = NO_2^--N concentration, g/m^3
NO_3-N = NO_3^--N concentration, g/m^3
$(NH_4\text{-}N)_e$ = concentration of NH_4-N in the effluent, g/m^3
$(NH_4\text{-}N)_{in}$ = concentration of NH_4-N in the influent, g/m^3
NOx-N = the amount of nitrogen oxidized to nitrate, g/m^3
$P_{,bio}$ = biomass production, kg VSS/d
P_{TSS} = production of TSS, kg TSS/d
P_{VSS} = production of VSS, kg VSS/d
Q = flowrate, g/m^3
Q_r = returned sludge flow rate, g/m^3
$q_{D,20}$ = specific denitrification rate at 20°C, g NO_3^- removed/g VSS h
$q_{D,T}$ = specific denitrification rate at temperature T(°C), g NO_3^- removed/g VSS h
q_N = specific ammonium-N removal rate, g NH_4^+-N removed/g VSS d
$q_{N,max}$ = maximum specific ammonium-nitrogen removal rate, g NH_4^+-N removed/g VSS d
q_{NO3} = specific nitrate removal rate, g NO_3^- removed/g VSS d
$q_{NO3,max}$ = maximum specific nitrate removal rate, g NO_3^- removed/g VSS d

R = returned activated sludge recycle ratio
R_o = oxygen demand, kg/d
S = growth-limiting substrate concentration (ammonium for ammonium oxidizer and nitrite for nitrite oxidizer), g/m^3
S_C = concentration of organic substrate, g/m^3
S_{NO3} = nitrate-nitrogen concentration, g/m^3
SRT = solids retention time, d
T = temperature, °C
TSS = total suspended solids, g/m^3
TSS_{in} = total suspended solids in the influent, g/m^3
V = the aeration tank volume, m^3
VSS = volatile suspended solids, g/m^3
VSS_{in} = volatile suspended solids in the influent, g/m^3
$(VSS)_{nb}$ = nonbiodegradable VSS, g/m^3
$X_{r,TSS}$ = TSS concentration in the returned activated sludge, g/m^3
X_{TSS} = TSS concentration in the aeration tank, g/m^3
X_{VSS} = VSS concentration in the aeration tank, g/m^3
Y = biomass yield, g VSS/g biodegradable COD
Y_n = biomass yield for nitrifying bacteria, g VSS/g NH_4-N
Y_N = biomass yield for ammonium oxidizer, g VSS/g NH_4-N
Y_{TSS} = Observed growth yield based on TSS, g TSS/g BOD
Y_{VSS} = Observed growth yield based on VSS, g VSS/ g BOD
θ = Simplified Arrenhius temperature-dependent constant

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14
Anaerobic Digestion

Jerry R. Taricska, David A. Long, J. Paul Chen, Yung-Tse Hung, and Shuai-Wen Zou

CONTENTS

Abstract Anaerobic digestion results in the biological breakdown of the sludge into methane, carbon dioxide, unusable immediate organics and a small amount of cellular protoplasm, under anaerobic conditions, mainly for sludge stabilization and volume reduction. This chapter introduces anaerobic digestion theory, biochemistry, microbiology, organic loading, temperature control, digester design, gas collection, and methane gas use, maintenance, and design examples.

Key Words Anaerobic digestion • digester design • methane production • gas collection • gas use • sludge pumping • sludge stabilization • volume reduction.

1. INTRODUCTION

Conversion of the organic material in solid wastes to methane-containing gases can be accomplished in a number of ways, including hydrogasification, pyrolysis, and anaerobic digestion. Hydrogasification is usually associated with the conversion of petrochemical raw materials. Although the process has been tried with solid wastes, it is not well defined and

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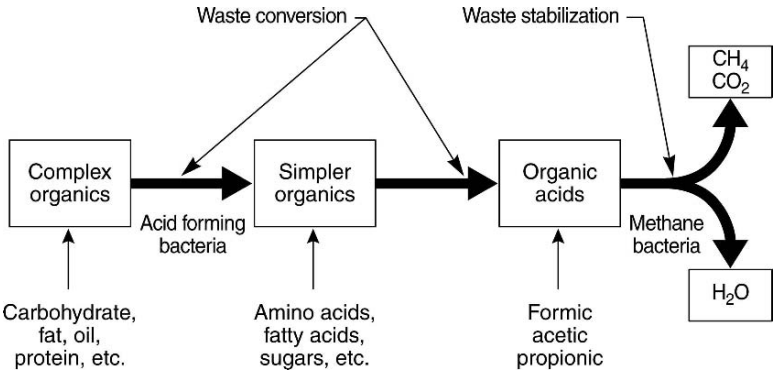

Fig. 14.1. General anaerobic biological reactions.

therefore is not considered in this book. The production of methane from solid wastes by pyrolysis has been considered previously. The production of methane from solid wastes by anaerobic digestion, or anaerobic fermentation as it is often called, is described in the following discussion.

Anaerobic digestion refers to the anaerobic decomposition of organic matter, resulting in partial gasification, liquefaction, and mineralization. The process is generally considered to be a two-stage biological process involving waste conversion and stabilization (Figure 14.1). The end products are principally methane (CH_4), carbon dioxide (CO_2), and stable organic residues.

Anaerobic digestion of solid waste and/or wastewater sludge has long been used to stabilize organic wastes before final disposal of these wastes. Among the benefits to be realized from such treatment are:

a. A reduction in organic content of the sludge.
b. Improved sludge dewaterability.
c. Destruction of most pathogens.
d. Generation of a potentially valuable byproduct (methane).
e. Volume reduction.

In addition to anaerobic digestion of solid waste and/or sludge, anaerobic treatment of wastewaters (particularly certain industrial wastes) has been receiving added attention in recent years. In addition to methane production, advantages cited for the anaerobic treatment process are as follows:

a. A high degree of waste stabilization may be obtained.
b. Relatively small amounts of residual organic waste are produced.
c. No oxygen is required.
d. Nutrient requirements are low (1).

Not only is interest in anaerobic processes being generated because of their waste treatment potential, but in our increasingly energy conscious society, the potential for generating methane from waste materials takes on added significance. In order for the process

potential to be fully realized, both design and operational fundamentals must be properly addressed.

The purpose of this chapter is to familiarize the reader with the theory of anaerobic processes and to present currently accepted design practices. Some attention also will be given to operational considerations as they affect selection and design of anaerobic unit processes.

2. THEORY

2.1. Nature of Organic Wastes

Where solid wastes are to be digested, special preparation of the solid wastes before digestion is necessary. The solid wastes first should be sorted and shredded to a size that will not interfere with the proper functioning of digester equipment and transport systems. It normally is necessary to add moisture and nutrients (pH adjustment may be required) to form a slurry that can be heated before feeding the mixture into the digester. Sewage sludge often is used to provide the necessary moisture and nutrients.

The majority of the sludges that are of concern in the design of anaerobic digestion facilities are of municipal wastewater origin. These sludges result from the settling out of solids in the sedimentation processes and may or may not contain biological waste solids from secondary treatment and chemical sludges from advanced waste treatment processes. Industrial wastes that are tributary to publicly owned treatment works may contain inorganic and organic solids that can alter the characteristics of what would otherwise be "typical" municipal wastewater sludges. An estimate of the amount of sludge to be expected from primary settling of raw wastes can be made by use of per capita estimates of suspended solids concentration together with estimates of the primary tank efficiency and the percentage total solids (TS) concentration in the settling tank underflow. If data are available on raw wastewater and primary clarifier effluent suspended solids concentrations together with appropriate flow data, estimates of the sludge mass and volume also can be made. Both of these techniques will be illustrated in the design examples presented in a later section of this chapter.

Solid waste and/or wastewater sludge to be digested normally are characterized on the basis of the percentage total solids (TS) and volatile solids (VS) fraction expressed as a percentage of the TS, both on a dry weight basis. It would be wise for the reader to review the analytical procedures for the laboratory determinations for residue in Standard Methods (2). Examination of organic wastes for other specific constituents, heavy metals, and so on, may be desirable where industrial wastes are present to ensure that such materials will not interfere with the anaerobic digestion process or limit the disposal of digested solids.

2.2. Biochemistry and Microbiology of the Anaerobic Process

As was indicated earlier, the anaerobic digestion process is considered to be a minimum of a two-stage biological reaction involving at least two different groups of microorganisms, acid-forming bacteria (saprophytic) and methane-forming bacteria. Complete understanding of all of the metabolic pathways, organisms involved, and so on is still lacking, but the general reactions involved have been identified.

2.2.1. Acid Phase

The first stage of the two-stage anaerobic process generally is considered to include the conversion of complex organic compounds into simpler organic compounds and finally into the organic acids, principally acetic acid (CH_3COOH) by acid-forming bacteria. Acetic and propionic acids are the most important volatile acids frequently occurring in "sick" digesters. As would be expected, little actual stabilization of organic wastes occurs during the first stage because the complex organic compounds (fats, carbohydrates, and proteins) are merely changed into simpler organic compounds.

The status of knowledge concerning the acid phase microbiology is not as far developed as is the microbiology of methane production. However, through methods similar to those used for ruminant bacteriology, it has been shown that most of the organisms responsible for acid production are obligate anaerobes (3). This fact means that it is essential to exclude oxygen from the digester environment to avoid oxygen toxicity. In order for the complex organics to be used by the acid-forming bacteria, they first must be solubilized by enzymatic action. The bacteria produce the necessary extracellular enzymes to accomplish this reaction.

2.2.2. Methane Fermentation Phase

Two basic mechanisms have been identified as biochemical pathways for the production of methane from the end products formed during the first (acid) stage (4). Of the two reactions shown below, Eq. (2) generally is considered to represent the more important reaction.

$$CO_2 + 8\,H^+ \xrightarrow{\text{Methane-forming bacteria}} CH_4 + 2\,H_2O \tag{1}$$

$$CH_3COOH \xrightarrow{\text{Methane-forming bacteria}} CH_4 + CO_2 \tag{2}$$

As is the case with all anaerobic bacteria of importance in wastewater treatment, the methane formers are very slow growing and may be subject to "wash out" at short hydraulic retention times unless recycling of microorganism is used. Reactor configuration will be considered in more detail in the next section.

The methane formation step is where the major waste stabilization occurs. Although, the methane formed is a high energy compound, its potential recovery and use makes it possible to show waste stabilization efficiency for the anaerobic process that is quite comparable to that expected from conventional aerobic treatment processes.

The methane formers do not constitute a single group (genus or species) of bacteria. Hence, a change in the predominance of the acid-forming bacteria present in a particular digester because of a change in substrate, and so on, may cause an upset, through the production of new first-stage end products that then require a group of methane organisms that are not present in sufficient numbers for a balanced condition to exist. The necessary balance should reestablish itself unless the influent waste or sludge characteristics or other factors are constantly changing. Figure 14.2 summarizes the major metabolic pathways involved in the anaerobic digestion process.

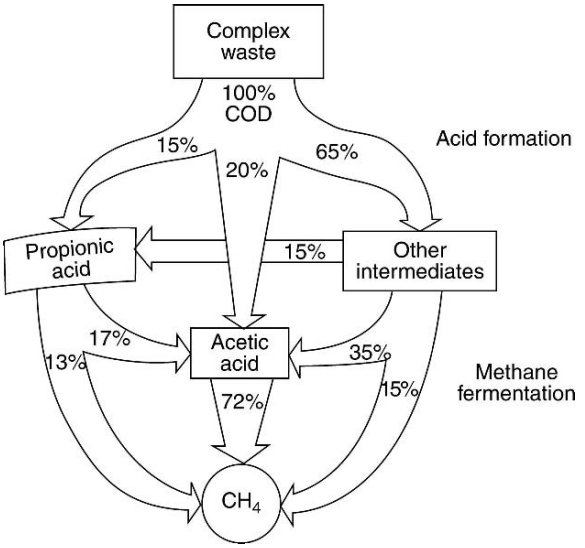

Fig. 14.2. Metabolic pathways in complex waste stabilization by anaerobic processes. After (1).

2.3. Reactor Configurations

There are four basic reactor configurations that have been used for the design of anaerobic unit processes. The four are:

a. Single-stage, unmixed.
b. Two-stage, mixed primary.
c. Anaerobic contact process with sludge recycle.
d. Anaerobic filter.

Of these four, the first two normally are used for digestion of solid wastes and/or wastewater sludge. Figure 14.3 illustrates the basic reactor types for so called conventional (standard rate) and high rate sludge digesters.

The majority of digestion tanks have circular cross sections. The conventional digester is unmixed and normally would be designed as a single-stage unit that allows digestion, supernatant separation and withdrawal, and stabilization and withdrawal of concentrated sludge to be accomplished in a single tank.

High-rate digestion systems are designed usually as two-stage systems wherein the primary stabilization of sludge is accomplished-in the first stage using a mixed digester. Separating supernatant from concentrated sludge and gas storage are provided for in the unmixed second-stage digester. The term "stages" as used in conventional engineering practice does not refer to the biochemical stages of acid production and methane formation discussed above, but rather to the physical processes of mixing and sludge-supernatant separation. Some studies (5) have suggested the design of digesters based on separation of the two biochemical stages of digestion, but this concept has not yet been accepted as general practice.

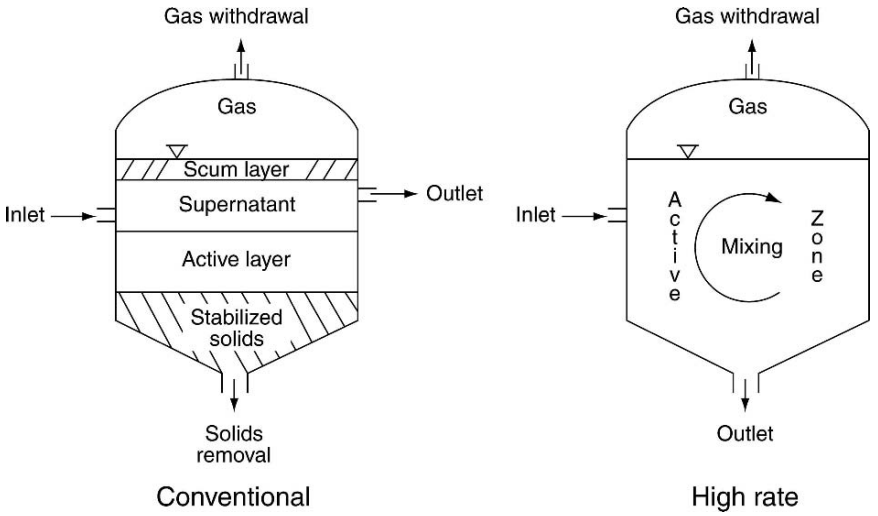

Fig. 14.3. Digester reactor configuration.

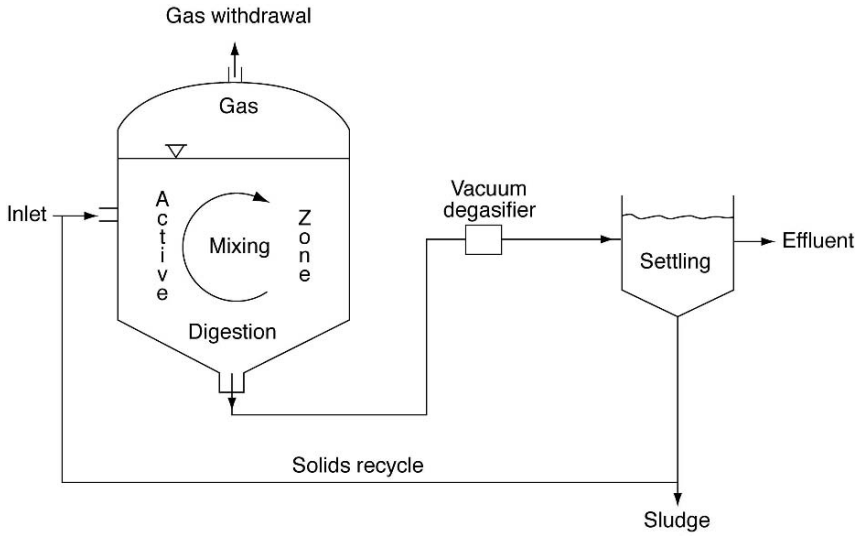

Fig. 14.4. Anaerobic contact process schematic.

The advantages of mixing the reactor to optimize the biological reactions involved should be readily apparent to the reader. In an effort to develop anaerobic technology to the point where it could be applied to the treatment of total wastes rather than just sludges, the anaerobic contact process (Figure 14.4) and the anaerobic filter have been developed (6, 7). The anaerobic contact process is simply an anaerobic activated sludge process and is designed and operated in much the same manner as complete mixed activated sludge (CMAS) process. The

major difference in the two, other than the obvious oxygen relationships is that some means of degasifying the reactor mixed liquor to permit gravity separation of the solids for recycle must be employed in the anaerobic contact process. The low growth rate in the anaerobic system makes solids recovery and recycling critical considerations for the satisfactory operation of such systems. The process has been used to treat certain industrial wastes that normally have high temperatures and high organic composition. The successful treatment of packing house wastes has demonstrated the suitability of the process for such wastes (8).

The anaerobic filter is basically an anaerobic trickling filter. The attached growth feature in this unit process may facilitate and enhance the retaining of the biomass needed to effect waste stabilization. This process has been employed less commonly today, compared to the anaerobic contact process, but it may find application in specific instances.

2.4. Organic Loading Parameters

Since the anaerobic process is a biological treatment process, loading parameters are most meaningful if they are expressed in terms of organic loadings. For solid wastes and organic sludges; loadings most commonly are based on volatile solids (VS), whereas for dilute wastewaters loadings would be expressed in terms of biochemical oxygen demand (BOD) or chemical oxygen demand (COD), as is the case with aerobic unit processes. This chapter emphasizes only anaerobic digestion of solid wastes and organic sludges. Conventional environmental engineering practice has been to express digester loadings on a weight to volume basis per unit time, normally, kilograms of VS per day per cubic meter of volume ($kg/d/m^3$).

The stability of the anaerobic process and the rate of gas production are both dependent upon organic loading rates. At higher loadings, the process often becomes unbalanced because of the excessive production of volatile acids. Carbon dioxide production under these conditions often will cause foaming of the digester and contribute to operating problems. Maintenance of uniform or near uniform loading rates based on frequent, or if possible, continuous additions of raw sludge to the digester will yield the most consistent digester operation.

The two basic modes of operation of sludge digesters generally are differentiated in terms of organic loading rates. The standard rate (unmixed) digester usually is loaded in the range of 0.48 to 1.6 kg $VS/d/m^3$ (0.03 to 0.10 lb $VS/d/ft^3$), whereas the high-rate digester (mixed) normally would be loaded in the range of 1.6 to 6.4 kg $VS/d/m^3$ (0. 10 to 0.40 lb $VS/d/ft^3$). Design loadings will be discussed in more detail in the section on Anaerobic Reactor Design and Sizing below. The degree of stabilization of sludges by anaerobic digestion is best expressed in terms of the volatile solids reduction that has occurred during digestion. Figure 14.5 illustrates accepted practice in determining when sludge is considered to be stabilized.

As indicated earlier, solid waste-sewage sludge mixtures have been digested successfully. Although mixtures containing from 50% to more than 90% solid wastes have been used, a 60% mixture appears to be a reasonable compromise. Loading rates for solid waste-sludge mixtures are not well-defined at present, but rates in the range of 0.64 to 1.6 kg $VS/d/m^3$ (0.04 to 0.10 lb $VS/d/ft^3$) should be satisfactory. Volatile solids reduction in such mixtures ranges

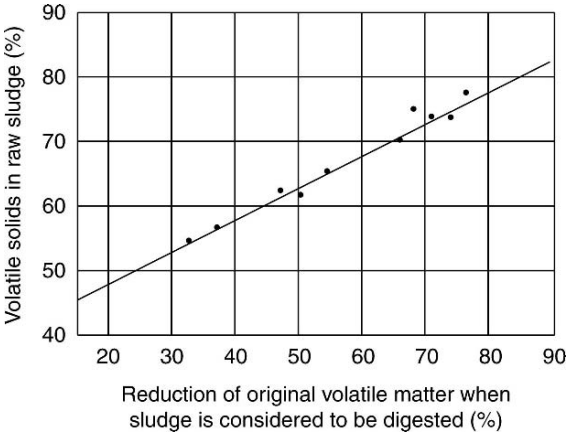

Fig. 14.5. Reduction in volatile matter by digestion.

from 60% to 80%, depending upon the amount of inert material present in the undigested mixture.

2.5. Time and Temperature Relationships

As with all biological systems, time and temperature are important factors in determining the degree of organic waste stabilization to be obtained. Sludge digestion in unheated digesters represents outdated technology except under special circumstances. Two general ranges of temperature have been investigated and used for sludge digestion in current practice. Mesophilic digestion at temperatures ranging from 30° to 37.5°C (86° to 100°F) is most commonly employed with the majority of installations operating at approximately 35°C (95°F). Thermophilic digestion at temperatures ranging from 48° to 57°C (120° to 135°F) has been used on a limited basis.

Generally biological reactions follow the Arrhenius principle of chemical reaction rates. Figure 14.6 shows the general relationship between temperature and digestion efficiency, based on solids retention time in days. As can be seen from Figure 14.6, digestion at elevated temperatures in the thermophilic range is more efficient in terms of the biological reactions involved (9). However, when the additional heat requirements necessary to operate at the higher temperatures are considered, it seldom is cost effective to operate in the thermophilic range.

The time required to obtain a desired degree of waste stabilization is primarily a function of temperature and of mixing, for reasons that have been stated above. In addition to these parameters, volatile solids reduction (waste stabilization) has been shown to be related to the raw sludge characteristics, as shown in Figures 14.7 and 14.8. Conventional digesters (standard rate) normally are designed with detention times of 30 to 60 days, whereas high-rate systems are designed with detention times of approximately 15 days in the first-stage mixed digester (10).

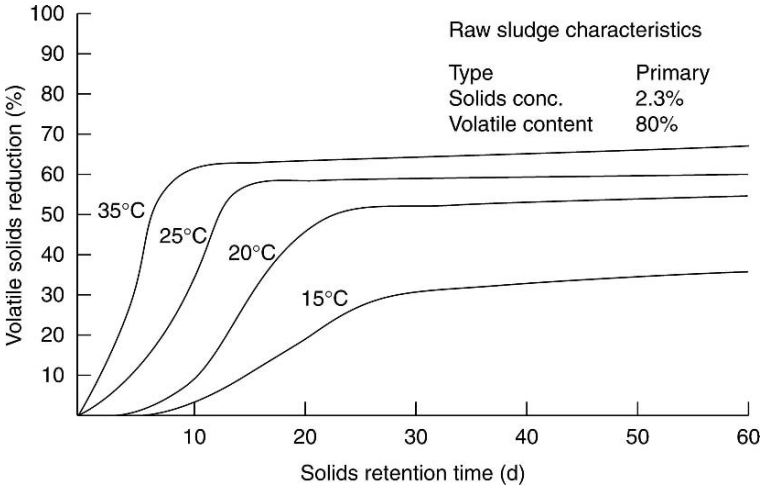

Fig. 14.6. Effect of solids retention time and temperature on volatile solids reduction in a laboratory scale anaerobic digester (19).

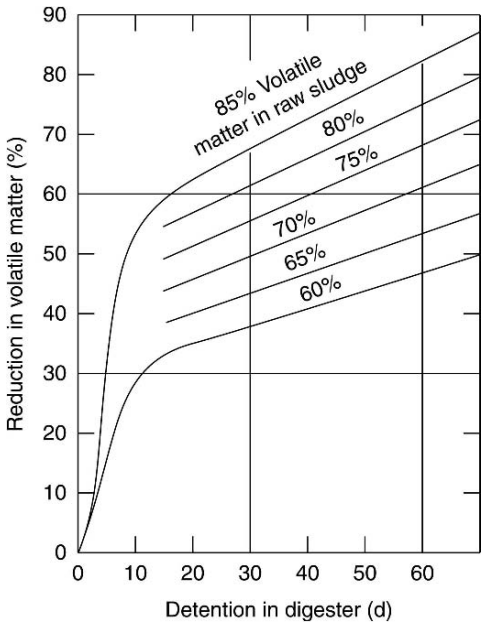

Fig. 14.7. Reduction in VS in raw sludge (19).

2.6. Nutrient Requirements

Nitrogen and phosphorus deficiencies are the two most commonly encountered inorganic nutrient deficiencies in wastewater treatment. These nutrients usually are present in sufficient amounts in municipal wastewaters and in sludges derived from treatment of municipal wastes,

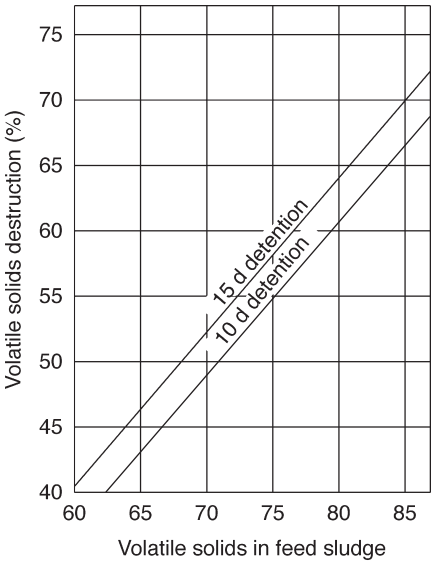

Fig. 14.8. Expected volatile solids destruction during high-rate digestion (19).

but they may not be present in the required amounts in all industrial wastes. Analyses of industrial wastes should include these two parameters. Nutrient requirements normally are approximately 11% and 2% of the weight of biological solids weight processed for nitrogen and phosphorus, respectively (1). Pilot studies can be helpful in assessing nutrient requirements for a particular system.

2.7. Gas Production and Use

The principal gases produced during the anaerobic digestion process are methane and carbon dioxide. Small amounts of hydrogen sulfide (H_2S) also normally are present and it may be noticeable in terms of the odor characteristics of the digester gas. Gases from a properly operating-digester should contain approximately 65% methane and 35% carbon dioxide by volume, with only traces of other gases. During digester upsets, the percentage of carbon dioxide in the off gases will increase. The fuel value of digester gas is approximately $5{,}850\,\text{kg-cal/m}^3$ ($656\,\text{BTU/ft}^3$), although this value will vary depending upon the composition of a particular digester gas.

The amount of gas to be expected from digestion process is typically expressed in terms of: (a) volume of gas produced per unit weight of volatile solids added, or (b) volume of gas produced per unit weight of volatile matter destroyed during digestion. The values commonly used in practice for each method are: (a) 0.5 to $0.75\,\text{m}^3/\text{kg}$ (8 to $12\,\text{ft}^3/\text{lb}$) of volatile solids added, or (b) 0.75 to $1.1\,\text{m}^3/\text{kg}$ (12 to $18\,\text{ft}^3/\text{lb}$) of volatile solids destroyed. McCarty (1) has developed a theoretical relationship to estimate the methane production to be expected where the amount of organic matter added, the degree of waste use, and the amount of biological

solids produced are known. Equation (3) expresses this relationship in metric units.

$$C = 0.35(eF - 1.42A) \tag{3}$$

where C = methane produced, m^3/d; e = efficiency of waste use (0.80 to 0.95); F = organic material added, kg/d (ultimate biochemical oxygen demand BOD_L); A = volatile biological solids produced, kg/d. Use Equation 8 to determine A.

Digester gases have been used as fuel for:

a. Heating digesters and other treatment plant facilities.
b. Gas engines used to drive blowers.
c. Engine driven generators.

Storage of the gas, as well as particulate and moisture problems and H_2S contamination, have to be addressed properly to realize the full potential of this energy source.

3. DESIGN PRACTICE

3.1. Anaerobic Treatability Studies

Anaerobic treatability studies usually are not necessary before proceeding with design unless:

a. Mixed domestic and industrial wastes are to be treated and some possibility exists of toxic or inhibitory wastes being present in the organic residue to be digested.
b. Anaerobic contact treatment of an industrial waste appears to be feasible on the basis of wastewater characteristics, but design loading data, and so on, are nonexistent.

Studies in the first instance are necessary to ensure that digesters can be loaded within normal ranges without adverse toxic effects from industrial wastes. In the second case, the treatability study will define process loading and operating parameters for the full-scale installation. The use of "typical" loading parameters often is unsatisfactory for this application.

Treatability studies can be conducted either using batch or continuous flow reactors. The batch system is less complex to setup and operate and is more applicable to evaluation of sludges rather than for treatment of total waste stream. Continuous flow reactors with some form of solids recycle is advantageous for evaluating anaerobic contact treatment. A schematic diagram of a batch system reactor set-up for anaerobic treatability studies is shown in Figure 14.9.

For sludge digestion studies, the operating variables to be investigated are detention time and volatile solids loadings. Detention time is controlled by wasting a constant volume of material from the digester each day. Because digesters normally are operated as nonrecycle reactors, the hydraulic residence time (θ) and mean cell residence time (θ_c) or solids retention time (SRT) can be considered to be the same. Sludge additions can be made once a day, but twice daily is preferable. If possible, digester feeding schedules to be used in the full-scale installation should be simulated. Volatile solids loadings can be controlled by dilution or concentration of the sludge used in the study. The use of several digesters operating in parallel will enable the investigator to obtain comparative results over various operating ranges in the shortest possible time. The detailed study procedures are as follows:

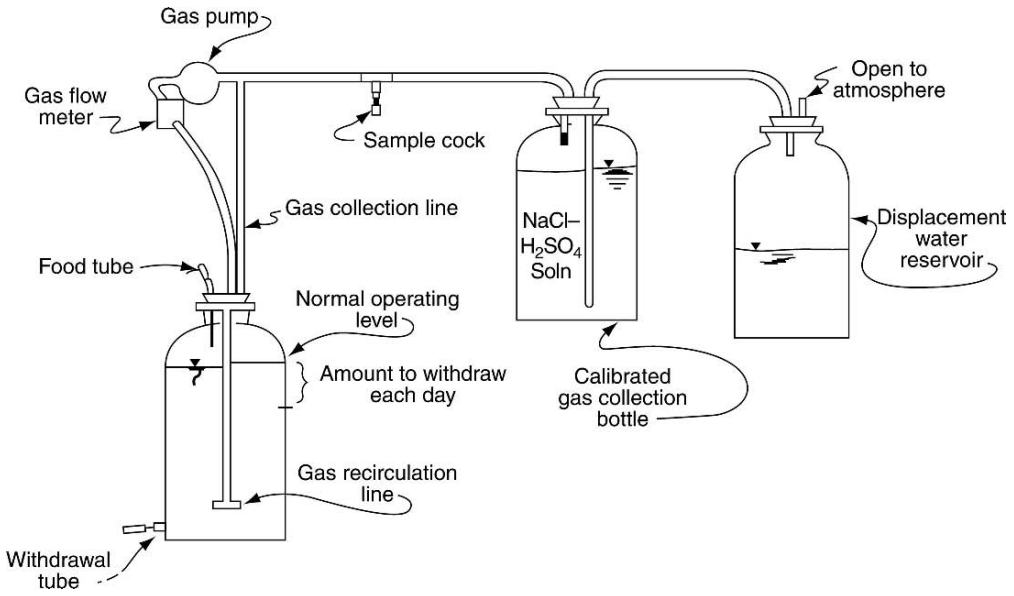

Fig. 14.9. Pilot digester setup for anaerobic process.

1. Seed the laboratory digesters using an actively digesting sludge from a municipal wastewater treatment plant. The seed sludge should be screened through 0.65 cm (1/4-in.) hardware cloth to remove, large particles that would clog feed and withdrawal lines.
2. Add the seed sludge to the digester and bring the digester contents to the normal operating volume by diluting with warm tap water. Care should be used to minimize the addition of air to the digester during all feeding and sludge withdrawal operations. Operational temperature of the digesters normally should be at 35 ± 1°C.
3. The reactor should be mixed continuously either by means of gas recirculation or mechanical mixers.
4. Feeding and withdrawal of sludge should not be started until gas production has been noted. Initial feedings should be such that full design loading is achieved over a period of several days.
5. Feed sludge should be screened and diluted with tap water or concentrated by removing supernatant to the proper total and volatile solids concentration for the volatile solids loading to be used at a particular detention time. Sludge can be made up ahead and stored at 4°C for up to 1 week.
6. Digesters can be fed once or twice daily; withdrawal of sludge normally would be done once daily after thorough manual mixing of the digester contents. Digested sludge grab samples should be taken daily and either analyzed daily or several days of grab samples combined into a composite sample and analyzed. Samples should be stored at 4°C.
7. Gas production can be measured by displacement of liquid from the calibrated gas collection bottle. Some error is introduced if the water levels in the gas collection bottle and the reservoir bottle are not equal, but the error should be small (less than 5%) in most instances.
8. Analyses to be performed include the following:
   a. Total and volatile solids in the raw and digested sludge (daily).
   b. pH (daily)

c. Volatile acids (3/wk)
d. Gas analysis (CH_4(%)CO_2 (%), 3/wk).
e. Alkalinity (daily).

9. Evaluation of data should include the following:
   a. Volatile solids loading and reduction under the various operating conditions.
   b. Gas production per unit weight of volatile solids added and destroyed.
   c. Plots of volatile solids loading, VS reduction, volatile acids, gas production versus time for each unit.

If the study is for the purpose of evaluating the anaerobic contact process for treatment of the total waste stream rather than for evaluation of sludge digestion, essentially the same procedure as outlined above should be followed. The major differences in the two procedures would be as follows:

a. Continuous flow operation should be approximated by adding a feed reservoir and feed pump to the reactor setup. Withdrawal of the necessary volume from the digester to control the hydraulic residence time (θ) still can be made once a day.
b. Sludge recycle can be approximated by returning solids removed from the volume withdrawn daily to maintain the hydraulic residence time. Inventories of solids in the system and removed from the system can be used to calculate the mean cell residence time (θ_c) for a particular set of operating conditions.
c. BOD, COD, and VSS determinations would be made three times per week to evaluate various loadings and removal efficiencies, solids production, kinetic coefficients, and so on; similar to the procedures used in activated sludge treatability studies.
d. Gas production should be expressed in terms of BOD and COD loadings and removals.

3.2. Anaerobic Reactor Design and Sizing

Current practice for the design of anaerobic sludge digesters normally involves the use of the so-called "rational basis of design," i.e., determination of digester capacity based on volatile solids (VS) loading, temperature, extent of mixing, and so on. In the absence of calculations that justify the basis of design, the GLUMRB Standards (14) require that the following minimum digester capacities:

a. Completely Mixed Systems: Completely mixed systems shall provide for intimate and effective mixing to prevent stratification and to assure homogeneity of digester content. The system may be loaded at a rate up to 80 lb of volatile solids per 1,000 ft^3 of volume per day (1.28 $kg/m^3/d$) in the active digestion units. When grit removal facilities are not provided, the reduction of digester volume caused by grit accumulation should be considered. (Complete mixing can be accomplished only with substantial energy input.)
b. Moderately Mixed Systems: For digestion systems where mixing is accomplished only by circulating sludge through an external heat exchanger, the system may be loaded at a rate up to 40 lb of volatile solids per 1,000 ft^3 of volume per day (0.64 $kg/m^3/d$) in the active digestion units. This loading may be modified upward or downward depending upon the degree of mixing provided.

As was indicated earlier, however, standard rate digesters have been designed on the basis of loadings ranging from 0.48 to 1.6 kg $VS/m^3/d$ (0.03 − 0.10 lb $VS/ft^3/d$) and high-rate digester loadings may range from 1.6 to 6.4 kg $VS/m^3/d$ (0.10 to 0.40 lb $VS/ft^3/d$).

Table 14.1
Effect of sludge concentration and hydraulic detention time on volatile solid loading factor[a, b]

	Volatile solids loading factor ($kg/m^3/d$)			
Sludge concentration, %	HRT = 10 d	HRT = 12 d	HRT = 15 d	HRT = 20 d
4	3.06	2.55	2.04	1.53
5	3.83	3.19	2.55	1.91
6	4.59	3.83	3.06	2.30
7	5.36	4.46	3.57	2.68
8	6.12	5.10	4.08	3.06
9	6.89	5.74	4.59	3.44
10	7.65	6.38	5.10	3.83

[a] Based on 75% volatile solids content of sludge and a specific gravity of 1.02 for sludge (concentration effects neglected).
[b] After ref. (12).
HRT = hydraulic retention time.

Table 14.1 shows volatile solids loading factors as a function of sludge concentration and hydraulic retention time in the digester. These loading factors can be used to size high rate digesters. If mixed organic and chemical sludges are to be digested, the volume of the digester must be increased over that calculated using the above loading factors to accommodate the greater volume of fixed solids reaching the digester. This adjustment can be made by multiplying the table values by the ratio of the actual percentage volatile solids in the sludge to be digested to the 75% VS sludge used as the basis for calculating tabular values.

Reactor design based on process kinetics should be the ultimate goal of the process engineer. The state of the art for the design of activated sludge treatment systems has advanced far beyond that of anaerobic systems in this regard. Only limited experience and data are available to assist the design engineer in the design of anaerobic systems in a similar manner.

As indicated earlier, the anaerobic contact process is essentially an anaerobic activated sludge process so the kinetic models developed for activated sludge systems can be applied directly to the design of anaerobic reactors. It is only necessary to evaluate the kinetic coefficients for substitution into the models. These coefficients can be evaluated in treatability studies as outlined above.

Development of mathematical models for describing biological wastewater treatment processes has been described elsewhere (12, 13). Lawrence (11) has presented the development of models for application of processes kinetics to the design of anaerobic processes. The working relationships for these models are summarized in Table 14.2. Use of the models is highly dependent upon the availability of kinetic coefficients so that pilot studies are essential as part of the design process. In the absence of kinetic coefficients, the designer may use the values listed in Table 14.3 for design. Examples of use of the models will be presented in a later section.

Table 14.2
Summary of steady-state relationships for completely mixed biological waste treatment process[a,b]

Characteristics	Without solids recycle	Eq. no.	With solids recycle	Eq. no.
Specific efficiency	$E_s = \frac{100(S_o - S_1)}{S_o}$	(4)	$E_s = \frac{100(S_o - S_1)}{S_o}$	(4)
Effluent waste concentration	$S_1 = \frac{UK_s}{k - U}$	(5)	$S_1 = \frac{UK_s}{k - U}$	(5)
Microorganism concentration in reactor	$X = \frac{Y(S_o - S_1)}{1 + k_d\theta_c}$	(6)	$X = \frac{Y(S_o - S_1)\theta_c}{1 + k_d\theta_c\theta}$	(7)
Excess microorganism production rate	$A = \frac{YQ(S_o - S_1)}{1 + k_d\theta_c}$	(8)	$A = \frac{YQ(S_o - S_1)}{1 + k_d\theta_c}$	(8)
Hydraulic retention time, V/Q	$\theta = \theta_c$	(9)	$\theta = \theta_c\,(1 + r - r(X_r/X))$	(10)
Solids retention times: General	$\frac{1}{\theta_c} = \frac{YkS_1}{K_s + S_1} - k_d$	(11)	$\frac{1}{\theta_c} = \frac{YkS_1}{K_s + S_1} - k_d$	(11)
Solids retention times: Limiting	$\frac{1}{\theta_c^m} = Yk - k_d;\ S_o \gg K_s$	(12)	$\frac{1}{\theta_c^m} = Yk - k_d;\ S_o \gg K_s$	(12)

[a] Nomenclature for the kinetic models is listed at the end of this chapter.
[b] After refs. (12, 13, 47, 48).

Table 14.3
Suggested mean cell residence times for use in the design of complete mix digesters[a]

Operating temperature °C	θ_c^m, d	θ_c suggested for design, d
18	11	28
24	8	20
30	6	14
35	4	10
40	4	10

[a] After ref. (12).

As anaerobic systems are low growth systems, mean cell residence times (θ_c) or solids retention time (SRT) must be long enough to avoid washout of the active microbial mass. Solids recycle in the anaerobic contact process helps to maintain the necessary active biomass in the reactor, while maintaining a desired short hydraulic retention time (θ).

Design parameters for anaerobic filters are not well-defined in current practice. Pilot studies should be conducted to determine organic and hydraulic loading rates, and so on.

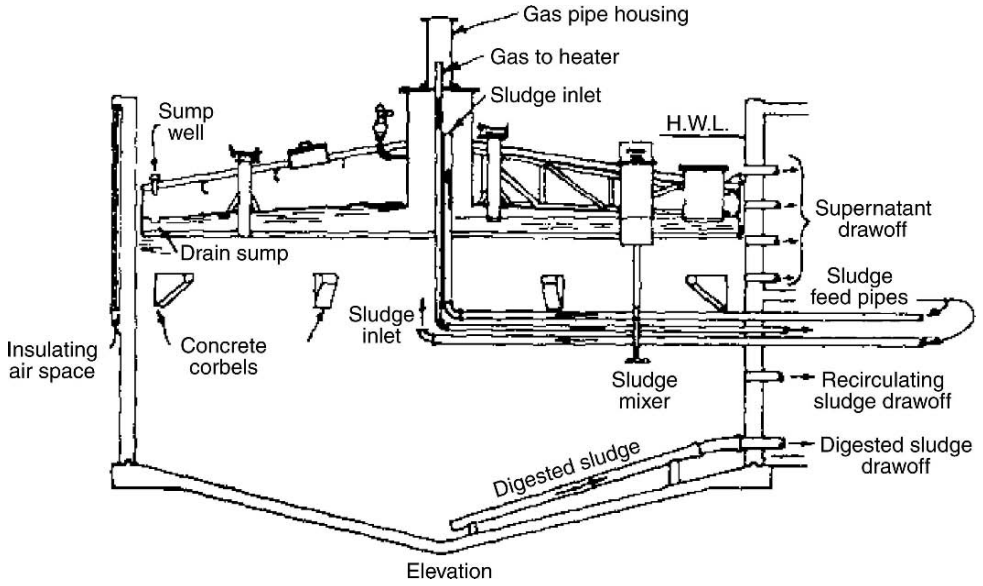

Fig. 14.10. Typical digester section (courtesy of Link-Belt).

3.3. Tank Construction and System Components

Most digesters are constructed of reinforced concrete and the most common configuration is that of a low vertical cylinder with a conical bottom (Figure 14.10). Tank diameters vary from approximately 6 to 34 m (20 to 110 ft) in increments to accommodate standard digester covers. Vertical side wall depths normally range from approximately 6 to 12 m (20 to 40 ft). The bottom slope should be at least 1 vertical to 4 horizontal when sludge is removed by gravity and reduced to 1 vertical to 12 horizontal when sludge is removed with suction mechanism (14).

Because of the necessity to heat digesters in most climates, digesters are insulated to reduce heat losses. Methods for insulating include the use of fiberglass or styrofoam panels, brick veneers with insulating air space, and mounded earth. Figure 14.10 illustrates the use of brick veneer with an insulating air space.

The use of multiple tanks is recommended wherever possible. This approach allows for operational flexibility, which can be critical during periods when digesters are under stress or when mechanical breakdowns occur. As a minimum, two tanks (usually of equal size) should be provided for high rate sludge digestion systems. The primary mixed digester normally is heated to provide an optimum environment for sludge stabilization and gas production. The second digester then serves as a solids separation tank and often as a gas holder.

The reactor for the anaerobic contact process is essentially the same as the primary digester described above. As indicated earlier, vacuum degasification (approximately 5.0 cm Hg of vacuum) of the digester effluent usually is necessary in the anaerobic contact system to achieve the good solids separation that is essential for the recycle of solids. Design of the sludge

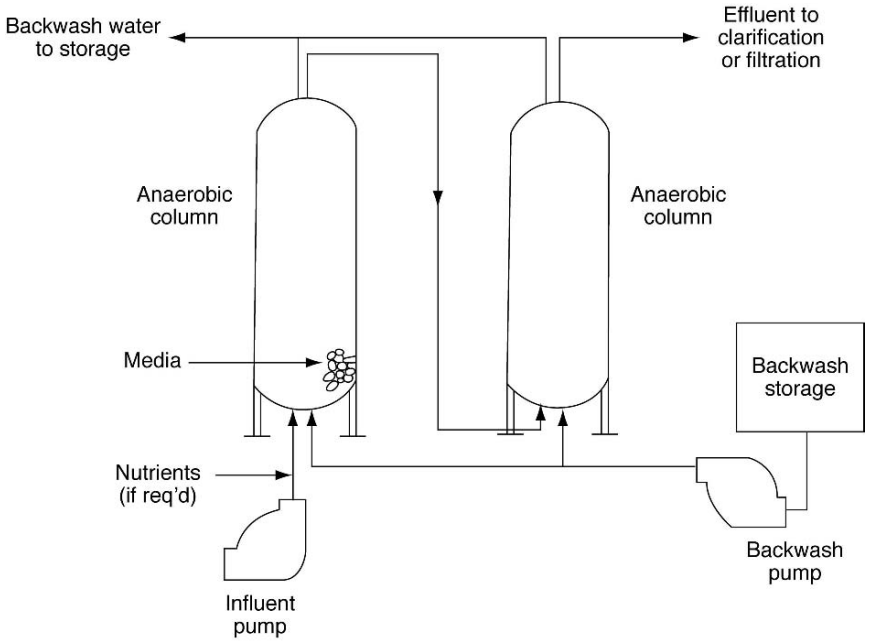

Fig. 14.11. Process schematic for anaerobic contact column.

separator is critical. Unfortunately, few data are available to assist the designer in selecting appropriate design criteria. Settling column tests of the degasified effluent should provide guidelines for selecting overflow rates and detention times for the full-scale installation. The use of the solids flux approach described by Dick and Young (15) should be investigated for applicability to the anaerobic contact process. Because the typical sludge generated in this process is a low density flocculent sludge, conventional plow-type clarifier equipment is not suitable for use in these systems. Suction-type sludge removal equipment should be specified for such applications. High rates of sludge return ($Q_r/Q \cong 3/1$) may be necessary because of the low solids concentration in the return sludge, so that pumping equipment must be selected with this flexibility in mind.

Packed columns or towers normally are used with the anaerobic filter system. The particular system configuration to be used will vary with the specific requirements of a particular installation. Figure 14.11 shows a typical process schematic for the anaerobic filter process. Various synthetic media are available for use as column packing. Some designs have combined filtration and anaerobic treatment in a single column. Columns may be operated either in upflow or downflow mode.

3.4. System Equipment and Appurtenances

A number of manufacturers produce equipment for use in anaerobic digestion systems. It is advisable to consult with them regarding application of their equipment for a particular installation.

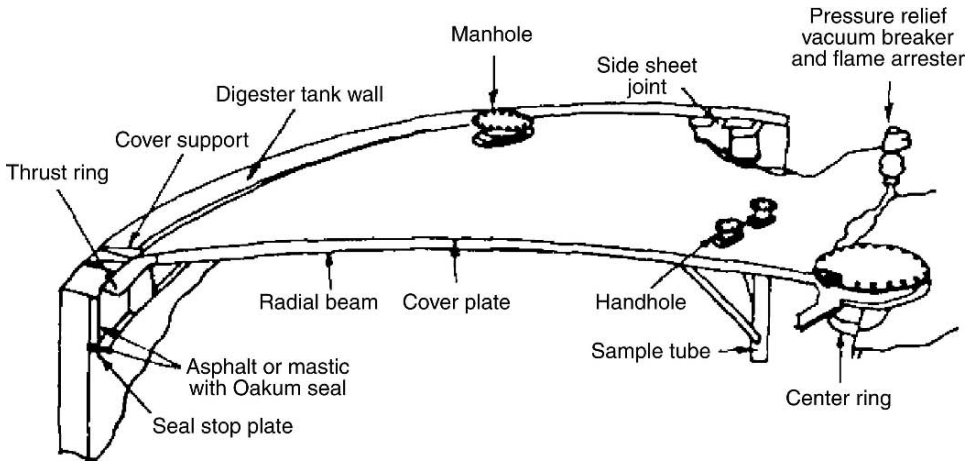

Fig. 14.12. Fixed digester cover and appurtenances (courtesy of Envirotech Corp., Emico BSP Div.).

3.4.1. Digester Covers

Digester covers normally are one of two types-fixed or floating. Primary digesters may be equipped with either type, but secondary digesters should be equipped with floating covers. Fixed covers ordinarily are constructed of reinforced concrete or steel, whereas floating covers usually are constructed of steel or steel framing with wood sheathing. Floating covers may float either on the liquid or gas in the digester. The gas holder cover is a floating cover designed essentially to float on the gas in the reactor. Floating covers offer more flexibility in operation of the digester because of the variable volume that is possible. The floating cover also minimizes the danger of mixing oxygen with the digester gas to form an explosive mixture (5.0% to 15.0% CH_4 in air by volume) and provides for gas storage under the gas holder. The advantages of floating covers generally offset their higher initial cost. Fixed cover digesters must be provided with a positive displacement feed and drawoff arrangement to avoid damage to the digester roof or the creation of dangerous explosive conditions. All tank covers should be equipped with vacuum and pressure relief valves and flame traps. A minimum of two manholes, 61 to 72 cm (24 to 28 in.) openings, should be provided in the digester cover. Gas-tight, quick-opening sample tubes also should be provided. Special precautions are necessary to prevent fire or explosions whenever digester covers are opened. Figures 14.12–14 illustrate the features of fixed, floating, and gas holder covers, respectively.

3.4.2. Mixing Devices

Effective mixing of the primary digester contents is essential to the proper operation of high-rate systems as indicated earlier. Various mixing systems are available for use, including various modifications of mechanical mixers and gas recirculation systems. Adequate digester mixing has not been defined specifically, but one manufacturer specifies that the mixing system should be adequate so that samples taken from various locations in the digester should not vary more than 10% in suspended solids concentration. Other recommendations suggest

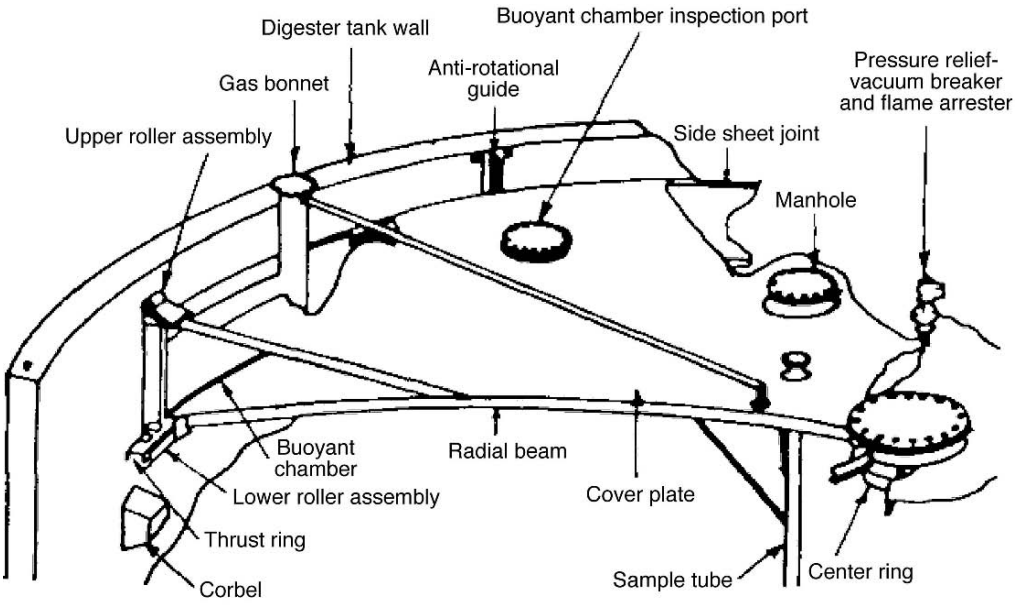

Fig. 14.13. Floating digester cover and appurtenances (courtesy of Envirotech Corp., Emico BSP Div.).

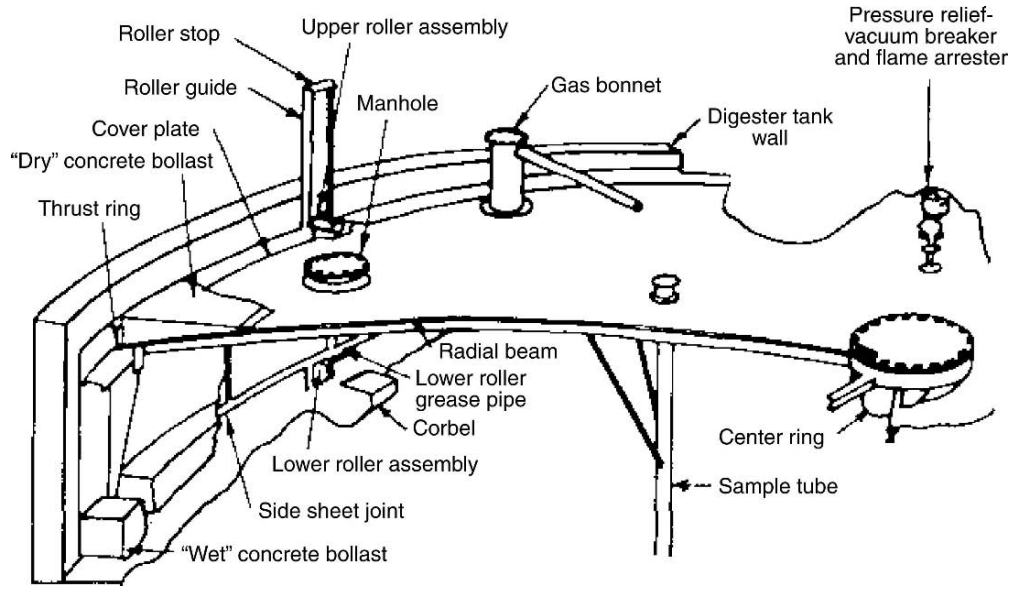

Fig. 14.14. Digester gas holder cover and appurtenances (courtesy of Envirotech Corp., Emico BSP Div.).

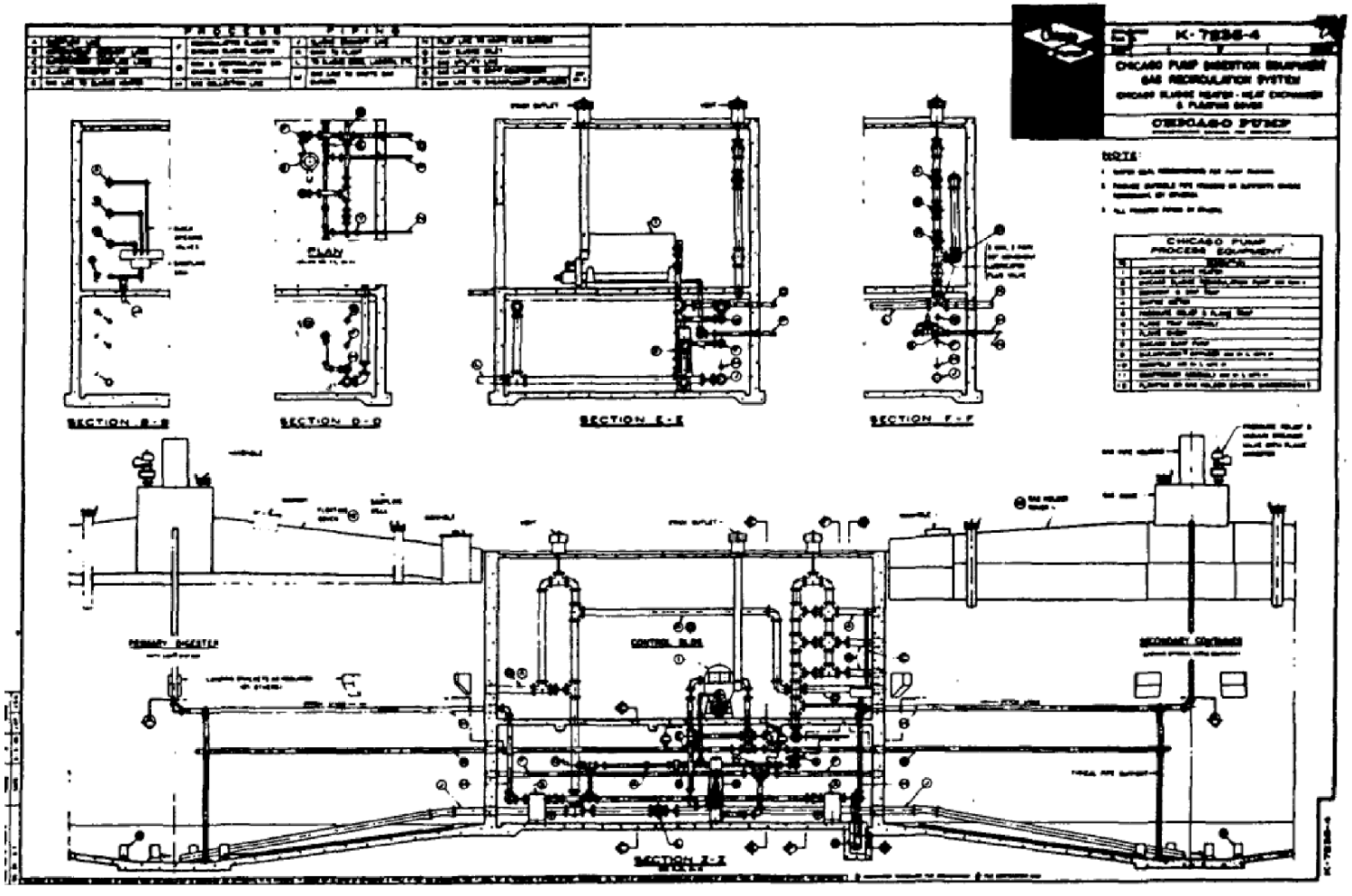

Fig. 14.15. Gas recirculation system (courtesy of Chicago Pump Co.).

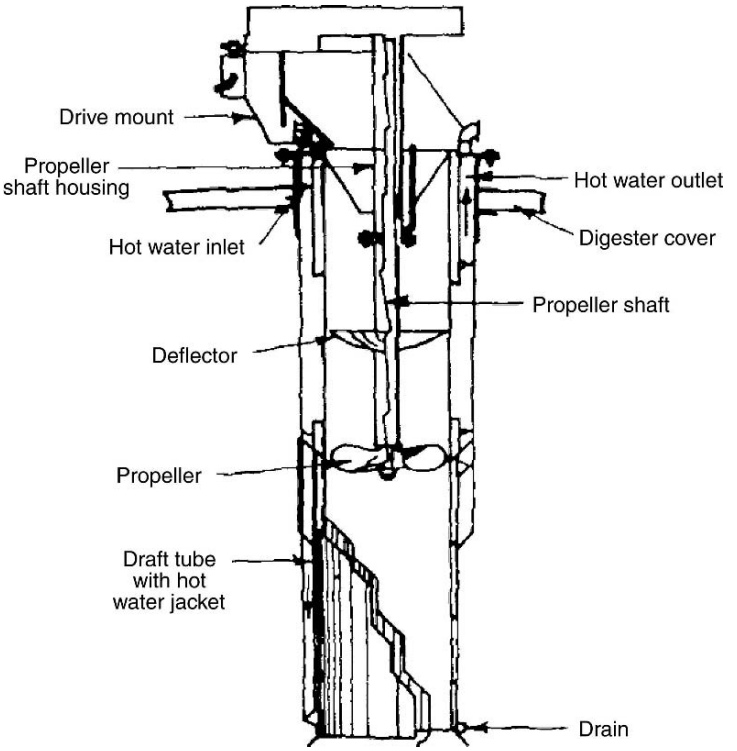

Fig. 14.16. Draft tube-type mixer (courtesy of Envirotech Corp., Emico BSP Div.).

that at least three turnovers per day of the entire digester contents be provided for in the recirculation system.

Gas recirculation systems have proven to be very popular in current practice. Several manufacturers furnish equipment that uses digester gas for mixing the digester contents. One of the newer developments in mixing technology is the aerohydraulic system developed by the Ralph B. Carter Company. Figure 14.15 shows a layout of a gas recirculation system. Draft tube-type mixers have been used in a number of installations. Various designs are available from the manufacturers and all are capable of providing the necessary degree of mixing. Figure 14.16 shows an example of one type of draft tube mixer. Mixing also can be provided by means of turbine-type mixers as shown in Figure 14.17.

Details on any of the mixing systems are available from the manufacturers.

3.4.3. Heating Systems

In most anaerobic digestion systems, it is necessary to supply heat from an external source to reach the desired operation temperature of approximately 35°C. A heat balance is necessary to determine the heat requirements for a given installation and to size the heating system components. The heat requirements include the amounts needed to heat the incoming raw sludge to the required temperature and to compensate for heat lost to the surrounding medium.

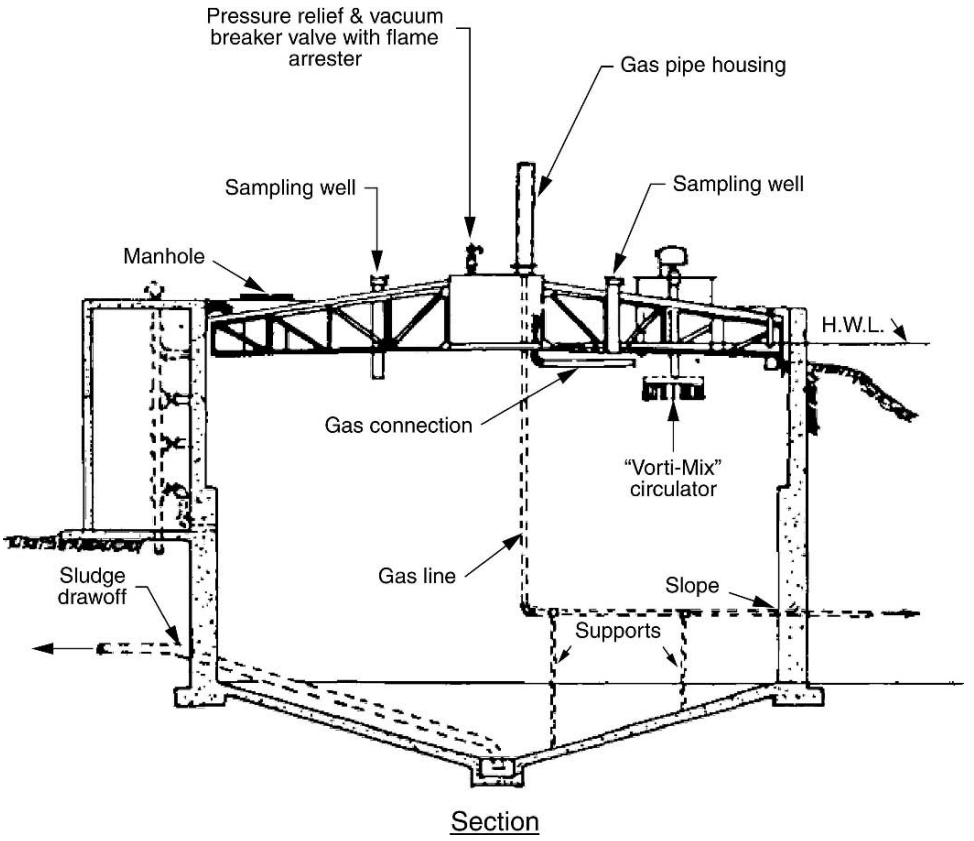

Fig. 14.17. Turbine-type mixing system (courtesy of Infilco/Degremont, Inc.).

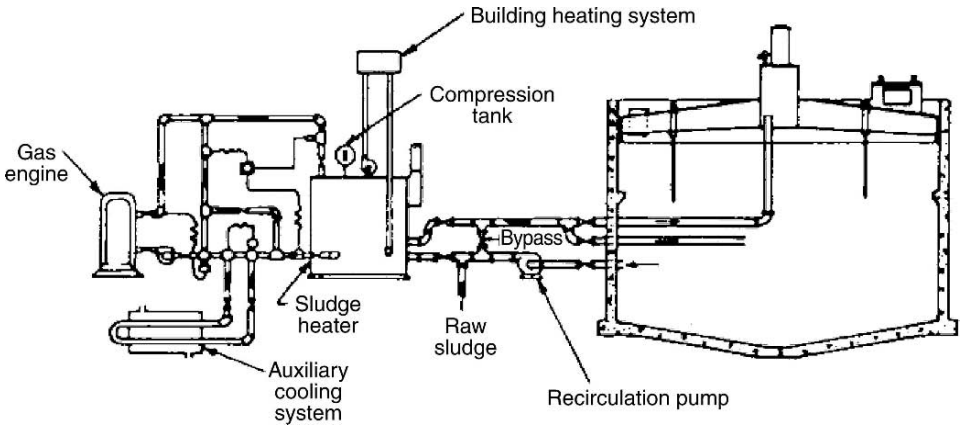

Fig. 14.18. Sludge heating system schematic (courtesy of Ralph B. Carter Co.).

Table 14.4
Digester heat transfer coefficients[a]

Digester section	$w/m^2/°C$	$BTU/h/ft^2/°F$
150 mm (6 in.) concrete roof	2.84	0.50
Floating cover with built-up insulated roof	1.36	0.24
300 mm (12 in.) concrete walls with air space insulation	1.99	0.35
300 mm (12 in.) concrete walls wet earth covered	1.42	0.25
300 mm (12 in.) concrete walls dry earth covered	1.02	0.18
Floor	0.68	0.12

[a]After ref. (18).

Figure 14.18 shows a schematic of a typical digester heating system. In two-stage systems, only the primary digester normally is heated.

The magnitude of heat losses from digesters is dependent upon the shape of the reactor and the type of construction used. Cylindrical digesters that have a diameter equal to depth are most efficient in terms of heat retention. Different materials of construction have different thermal transfer coefficients. Heat losses from digesters can be approximated from the following equation:

$$Q = UA(T_2 - T_1) \tag{13}$$

where Q = heat loss from the tank, w (BTU/h); U = heat transfer coefficient, $w/m^2/°C$ ($BTU/h/ft^2/°F$); A = surface area of tank element, $m^2(ft^2)$; T_1 = temperature outside the tank, °C(°F); T_2 = temperature inside the tank, °C(°F).

Overall heat transfer coefficients (U) are dependent upon the materials of construction, their relative conductivities and thicknesses, the degree of turbulence inside the tank, and the presence of earth or air outside the tank. The overall values of the heat transfer coefficients for different digester sections presented in Table 14.4 can be used to calculate heat losses with Equation (13). The temperature inside the tank (T_2) would be the normal operating temperature of the digester. The average ambient air temperature for the coldest 2-week period expected should be used for the temperature outside the tank (T_1).

An easier method of estimating digester heat losses that does not require the consideration of heat losses through each element of the digester has been used. Approximately $2{,}720\,w/100\,m^3$ ($2{,}600\,BTU/1{,}000\,ft^3/h$) will be lost from a well-insulated unmixed digester in the northernmost part of the United States. The values in Table 14.5 can be used to estimate digester heat losses for the conditions shown.

The amount of heat necessary to raise the temperature of the incoming raw sludge to the desired level can be calculated from the following:

$$H = WC(T_2 - T_1) \tag{14}$$

where H = amount of heat required, J (BTU); C = mean specific heat of raw sludge = $4{,}200\,J/kg/°C(1.0\,BTU/lb/°F)$; W = weight of sludge entering the tank per hour, kg (lb);

Table 14.5
Estimated heat losses from anaerobic digesters

	Heat losses					
	w/100 m^3			BTU/h/1000ft^3		
Digester conditions	Northern US	Middle US	Southern US	Northern US	Middle US	Southern US
Mixed and insulated	4190	2090	1260	4000	2000	1200
Mixed and noninsulated	5230	2620	1570	5000	2500	1500
Unmixed and insulated	2720	1360	840	2600	1300	800
Unmixed and noninsulated	4190	2090	1260	4000	2000	1200

T_2 = temperature of sludge in the tank, °C (°F); T_1 = temperature of raw sludge entering tank, °C (°F).

The average temperature of the raw wastewater during the coldest 2-week period of the year normally is used as the value for T_1.

Methods of heating anaerobic reactors include the use of external heat exchangers, jacketed draft tube mixers and internal pipe coils. The latter method generally is considered to be outdated technology and the method of choice normally would be use of external heat exchangers. Figure 14.19 illustrates a typical external heat exchanger of the type used for sludge heating applications. A major advantage of the use of such equipment is the ready access to the tubes for maintenance and cleaning. Recirculation of sludge through the exchanger also helps to mix the digester. The provision of multiple inlet and outlet points in the digester piping arrangement greatly facilitates operational flexibility and helps to better maintain the desired level of mixing.

An example of a jacketed draft tube mixer is illustrated in Figure 14.16. With this system it is necessary to provide an external boiler to heat the water for recirculation through the draft tube jacket. The circulation water temperature should be approximately 65.6°C (150°F).

Boilers and heat exchangers should be equipped for dual fuel use. Digester gas normally is used for fuel, but oil or gas should be available for use during startup or other periods when the digester gas production is insufficient to meet fuel needs. Indicating and recording thermometers should be provided to monitor the temperature of the incoming and return sludge and the hot water. Heating units should be sized to handle the heat requirements calculated above and may include building heat requirements where appropriate.

3.4.4. Gas Collection, Storage and Distribution

It is necessary to collect the gas that is generated during anaerobic digestion. The collected gas then can be used as a fuel source or burned (flared) to avoid creating a nuisance or potentially dangerous situation. As indicated earlier, gas-air mixtures must be avoided in the gas collection system to prevent explosion hazards. If the digester gas collection system is kept under positive pressure, air cannot be drawn into the system. Proper operation of digester cover systems will provide the needed positive pressure in the system.

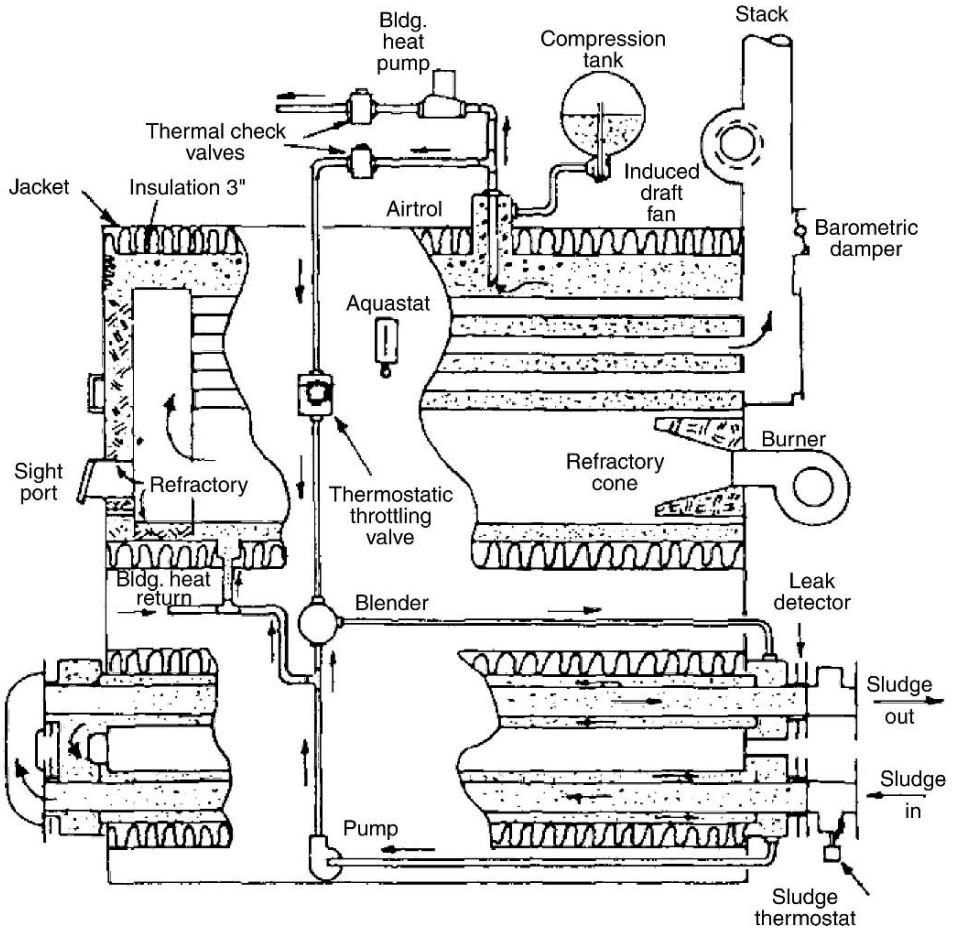

Fig. 14.19. External heat exchanger for use in anaerobic digesters (courtesy of Ralph B. Carter Co.).

Some gas storage normally is provided under the digester cover, as discussed previously. Storage is necessary to balance demand for gas used to fuel use equipment with gas production from the digesters. In addition to gravity type gas holders similar to floating digester covers, pressure-type holders also are used. Operating pressures used range from 1.4 to 7.0 kg/cm^2 (20 to 100 psi). Gas is pumped to the gas holder by means of a suitable gas compressor.

Gas collector and distribution lines must be sized properly to handle the maximum anticipated gas flows without excessive pressure drop. For systems with gas recirculation, the recycle gas flow must be taken into consideration in sizing gas lines. The maximum velocity in gas piping normally is limited to approximately 3.5 m/s (11.5 ft/s) to avoid high- pressure losses and carryover of moisture from condensate traps. Gas piping should be sloped a minimum of 1 cm/m (1/8 in./ft) with greater slopes, 2 cm/m (1/4 in./ft) where possible. Digester gas is very wet, so that drainage and removal of condensate from the gas system is important to proper operation.

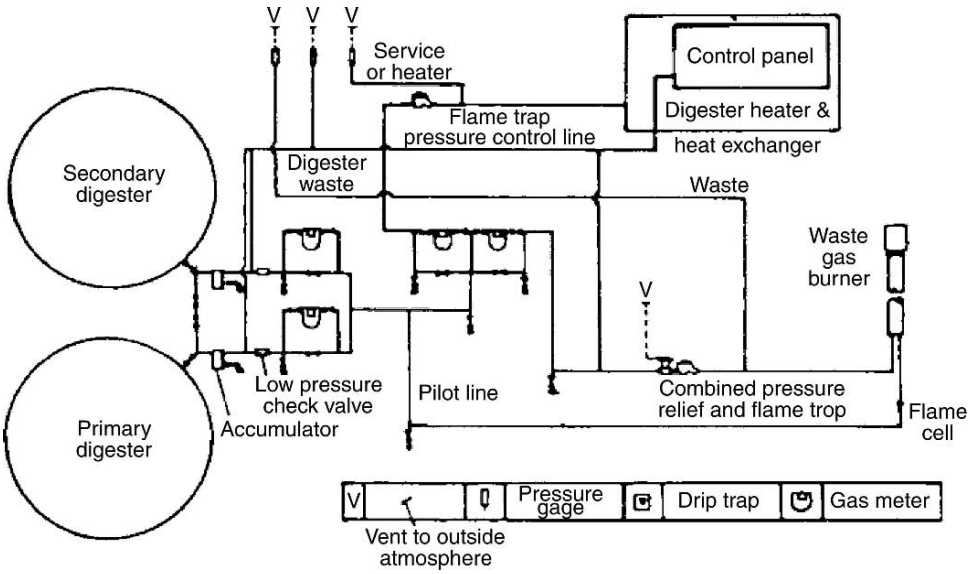

Fig. 14.20. Gas piping schematic of a modern anaerobic digestion system (courtesy of Envirotex).

In addition to the pressure and vacuum relief valves that are required as part of the digester cover appurtenances, flame traps, thermal valves, sediment traps and drip traps must be provided on all gas lines. Flame traps should be installed in all gas lines that connected to gas use equipment and should be placed as close as possible to the points of ignition. The use of thermal valves is recommended to provide additional protection against fire and explosion. Sediment traps are necessary to remove particulates carried over in the gas from the digester, scaling from corroding pipes, and other source of particulates. Manually operated drip traps should be located at all low points in the gas piping so that accumulated moisture can be removed before it impedes gas flow or causes damage to gas use equipment.

Accurate metering of gas produced, used, and wasted is essential to proper digester operation. Various types of gas flow meters, such as diaphragm, shunt flow, propeller, and so on, are available. Because digester gas is wet and dirty, selection of meter materials and construction that resists corrosion is of utmost importance. Bypass lines around the meters should be provided to facilitate removal of the meter for proper maintenance.

The use of manometers to indicate gas pressure in the system is desirable. Pressure regulators may be required at several points in the system depending upon the requirements of the gas use equipment. Design pressure in the gas system and digester cover normally is approximately 150 to 250 mm of water column (6 to 10 in.). Figure 14.20 shows a schematic of a complete digester gas system.

Waste gas burners (flares) generally should be provided to burn excess gas. The burner should be located at least 7.6 m (25 ft) away from any plant structure if placed at ground level. Burners may be placed on the control-building roof if it is located sufficiently far away from the digester tank. Safety considerations require that adequate ventilation be provided in all

enclosed areas where digester gas may accumulate. Electrical fixtures in these areas should comply with the National Fire Protection Association requirements for hazardous locations.

3.5. Gas Use

Use of digester gas is becoming increasingly more important as energy costs continue to rise. As pointed out in an earlier section, digester gas has a fuel value of approximately 5,850 kg-cal/m^3 (657 BTU/ft^3) and has been used in gas engines to drive pumps, blowers, and generators as well as for heating digesters and buildings.

Digester gas most commonly is used to fuel low-pressure hot-water boiler systems. Because of the potential for corrosion of vents and the resulting release of toxic and asphyxiating gases, digester gas should not be used as a fuel for "open flame" type unit heaters.

Use of digester gas as a primary fuel for driving dual-fueled reciprocating engines that are used as driving units has been practiced in a number of installations. It is possible to use the engine jacket water as a hot water source when engine use is continuous. Blending digester gas with a commercial fuel may be necessary to ensure continued deliverance of a fuel with suitable heat value to the engine. The required blending can be accomplished automatically in most dual-fueled engines.

Some use has been made of digester gas as a fuel for gas turbine drivers in recent years. However, these installations generally are much more expensive than those required for internal combustion engines and the additional costs may not be justified.

As digester gas is quite "wet and dirty," it is often necessary to install gas scrubbers to remove particulates and hydrogen sulfide. Removal of carbon dioxide will increase the heat value of the fuel, and equipment to accomplish this task may be justified in some instances. Hydrogen sulfide is a particular problem because it forms a corrosive liquid vapor when burned in combination with water vapor. The maintenance of boiler water temperatures above 82°C (180°F) will reduce the problem of corrosion caused by the condensing of the vapor in fire tubes and stacks. Proper preventative maintenance on gas engines is very important when using digester gas fuel because of the potential for corrosion, varnishing of cylinder walls, and so on, created by the presence of impurities in the gas. These problems are particularly severe when engine duty is not continuous. This gas is also malodorous and has rotten egg smell.

Hot water boilers and internal combustion gas engines generally require a gas pressure of 76 to 130 mm (3 to 5 in.) of water for proper operation and to ensure a positive pressure throughout the gas system. Gas turbine engines require a fuel pressure of 10.5 to 14.1 kg/cm^2 (150 to 200 psi), and thus compressors and high-pressure storage facilities are required for such installations in addition to gas scrubbing equipment.

3.6. Sludge Pumping and Piping Considerations

Proper operation of the sludge digestion system depends upon the ability to transport sludge in the system. Most systems require at least some pumping of sludge because sludge must be transfer from one sludge tank to other. The hydraulic characteristics of sludge can vary widely as a result of differences in viscosity, solids concentration, and so on, and the designer must take these factors into consideration in selecting pumps and piping to handle the sludge encountered. Centrifugal pumps (screw centrifugal or disc) normally are the most

economical for low viscosity sludge (waste activated sludge, dilute primary sludge, etc.), whereas positive displacement pumps (progressing cavity, peristaltic, etc.) should be selected for handling highly viscous sludge such as thickened primary sludge. Inlet and discharge pulsation dampeners should be considered when positive displacement pumps are used. As the sludge enters the dampeners, the trapped gas in the fluid (sludge) is compressed.

Head losses resulting from pipe friction can be significantly higher than those expected for water, so the designer must adjust head loss calculations accordingly. Brisbin (16) and Chou (17) investigated the flow of wastewater sludge in pipes and found that the Hazen-Williams C value varies as a function of sludge moisture concentration. They have recommended procedures to use for hydraulic calculations for sludge piping. Additional information on the hydraulic characteristics of sludge can be found in other design references (18, 19).

Piping flexibility is extremely important to permit proper management of sludge and supernatant in the digester system. Scale models of the digester piping, including valves, and so on, can be invaluable in the preparation of detailed plans for the digester installation.

4. MANAGEMENT OF DIGESTION

4.1. Control of Sludge Feed

Proper control of the raw sludge feed to the digester probably is the single most important process control mechanism available to the treatment plant operator. Volatile solids loadings should be controlled at optimum levels and additions of sludge should be made on a frequent regular basis rather than as the infrequent large additions that occur when sludge is pumped only once a day. The total solids concentration of the sludge to be added to the digester should be maintained as high as possible (consistent with handling considerations) to minimize the amount of water added to the digester. Time clock controls on the raw sludge pumps can be useful for this purpose, particularly when they are tied into a sludge density or sludge concentration meter. For smaller plants in which complete instrumentation of sludge pumping is not warranted, the operator should monitor sludge concentration through visual checks of the sludge being pumped or by other simple observations, such as of pump pressures, motor ammeter readings, the sound of the pump, and so on.

4.2. Withdrawal of Sludge and Supernatant

Although the quality control of supernatant withdrawals is not considered critical to process control of the digester, its potential impact on other treatment plant unit processes cannot be overemphasized. The return of poor quality supernatant to the plant head works or to other points is, in many instances, responsible for plant upsets and operating difficulties, particularly in aerobic treatment units. If the digester is being operated properly and sufficient flexibility in the withdrawal piping exists, the operator should be able to select supernatant of satisfactory quality (less than 5,000 to 7,500 TS mg/L). The use of chemicals, such as polymers, may be warranted in some cases to obtain satisfactory supernatant quality so as not to cause upsets in the plant. Preaeration of supernatant often is helpful where facilities permit such pretreatment.

Digested sludge withdrawal ordinarily does not create problems at most plants. Sufficient seed sludge should be left in the digester following withdrawal to maintain approximately 20 kg of actively digesting volatile solids in the digester for each 1 kg of raw volatile solids

added to the digester per day. The use of multiple sludge withdrawal points to permit selection of the best quality digested sludge and to avoid a buildup of grit, and so on, in the digester is recommended. The use of a solids inventory scheme to control additions and withdrawals of sludge and supernatant is invaluable as a process control tool for digestion process as well as for other unit processes in the solids handling train.

4.3. Maintenance of Reactor Stability

In addition to the proper control of sludge additions and withdrawals already discussed, proper temperature control and provision of adequate mixing are critical to maintaining reactor stability. Methods to be used for temperature control will depend upon the type of heating equipment provided in the digester installation. The particular temperature selected for operation of a given digester is not as critical as is the maintenance of a relatively constant temperature as long as it is in the normal range of 30° to 37.5°C (86° to 100°F). Temperature changes should be kept to no more than 0.5°C (1°F) per day if digester upsets are to be avoided. Temperature shocks to the digester can be minimized by heating the raw sludge before it is introduced into the digester. Heat requirements can be minimized by reducing the amount of water added to the digester with the sludge.

As indicated earlier, proper mixing of the entire contents of the digester is needed to optimize the biological reactions occurring in the digester. Good mixing prevents, or at least significantly reduces scum blanket formation, which results in more of the digester volume being available for sludge stabilization. The procedures to be used for mixing control in a particular installation also depend upon the type of equipment furnished in the digester.

4.4. Digester Performance Criteria

Although a number of different parameters can be used to monitor and control the anaerobic digestion process, the following are considered to be the most significant:

a. Volatile acids to alkalinity ratios.
b. Gas production and composition.
c. pH.
d. Volatile solids loadings and volatile solids reduction in the digester.

All of the above should be monitored on a regular basis because none of them alone give sufficient information about process conditions. Plotting the monitoring data is helpful, particularly in following the performance of digester, because the rate of change of the various parameters is more significant than are absolute numbers. This approach to process control makes it possible to detect indications of process upset as early as possible so that collective action can be instituted.

As with any process control, the importance of using proper sampling procedures that yield truly representative samples cannot be overemphasized. Selection of sampling locations, frequency of sampling and analysis, and so on must be adapted to meet the needs of a particular installation.

The operation and control of anaerobic digesters is discussed in more detail in the Water Pollution Control Federation MOP 11 (20) and MOP 16 (21), and the US EPA Operations Manual on Anaerobic Sludge Digestion (19, 46). The material contained in these publications

is extremely helpful to the designers of unit processes, and it should be reviewed early in the design stage of a project to design for optimal operability of the facility.

5. CAPITAL AND OPERATING COSTS

5.1. General

The information on costs presented in this section must be used with care. Cost data, which are included in Figure 14.21, represent average costs for capital construction and operation and maintenance (22). These data may be used for preliminary estimates for planning purposes and general comparisons among alternative treatment schemes. It must be pointed out that these data cannot be considered to be applicable in specific treatment plant estimates without further refinement and adjustment for local conditions.

5.2. Items Included in Cost Estimates

The cost data presented are based on a two-stage digestion system.

1. Capital Cost
   Capital costs include tanks, mixers, heating devices, controls, and all other appurtenances required for the process. Devices for the collection of gas from the digesters are included, but no provision is made for the use of this gas for power recovery.
2. Operating and Maintenance Costs
   Labor represents the most significant operating and maintenance cost for anaerobic digestion. This process requires a high degree of operating control and supervision for peak efficiency.

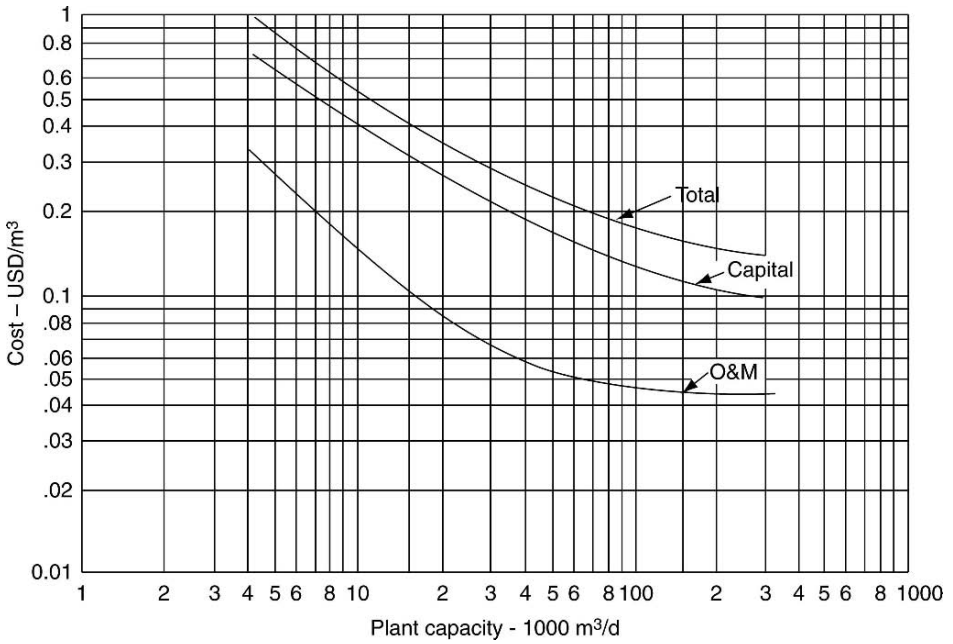

Fig. 14.21. Estimated costs of anaerobic digestion facilities (22).

Various tests must be run periodically to monitor the digestion process and make appropriate adjustments. Proper maintenance requires the cleaning of digesters periodically and repairing equipment.

The cost of final sludge disposal has not been included here, but must be considered for determination of total O&M costs.

The nature of the influent sludge will have some effect on the total costs of the sludge handling facility. The cost data presented above are based on an assumed municipal influent sludge from conventional sedimentation and biological processes. If chemical sludges are to be included, adjustments must be made in the overall sludge handling scheme to allow for the lesser reduction in solids that would occur in the anaerobic digester. These adjustments would have to be considered when estimating costs for the unit processes.

6. DESIGN EXAMPLES

The following examples are included to illustrate the design of anaerobic digesters in accordance with the procedures outlined previously.

6.1. Example Using Standards Design

Estimate the size of the two-stage digesters required to treat the sludge from a community of 40,000 persons. For the wastewater to be treated, it has been found that 0.10 kg/cap/d (0.22 lb/cap/d) of dry solids are contained in the raw wastewater. Primary settling removes 55% of the suspended solids originally present in the raw wastewater. Pilot studies of the secondary treatment processes have shown that 0.05 kg/cap/d (0.11 lb/cap/d) of waste sludge on a dry weight basis will be generated. Assume that the raw primary sludge contains about 5% total solids (95% moisture) and that the waste secondary sludge will be thickened to 4% total solids (96% moisture). The digested sludge should contain 8% total solids (92% moisture). All sludges are assumed to have a specific gravity of 1.02 and the raw sludge contains 75% volatile solids. Other pertinent design assumptions are as follows:

1. Only the primary digester will be mixed and heated to 35°C.
2. Sludge contains adequate nitrogen and phosphorus for biological growth.
3. The design loading is 1.28 kg VS/m^3/d in accordance with GLUMRB standards (14).

Solution

a. Compute the weight of volatile solids to be added to the digester daily.

$$\begin{aligned}
\text{Primary sludge (kg/d)} &= (40{,}000 \text{ persons})(0.10 \text{ kg/cap/d})(0.55) \\
&= 2{,}200 \text{ kg/d} \\
&= (2{,}200 \text{ kg/d})(2.2046 \text{ lb/kg}) = 4850 \text{ lb/d} \\
\text{Waste sludge (kg/d)} &= (40{,}000 \text{ persons})(0.05 \text{ kg/cap/d}) \\
&= 2{,}000 \text{ kg/d } (4{,}409 \text{ lb/d}) \\
\text{VS to digester (kg/d)} &= (2{,}200 + 2{,}000)(0.75) \\
&= 3{,}150 \text{ kg VS/d } (6{,}945 \text{ lb VS/d})
\end{aligned}$$

b. Compute the volume of raw sludge expected.

$$\text{Primary sludge}(\text{m}^3/\text{d}) = \frac{2200\,\text{kg/d}}{(0.05)(1.02)(1000\,\text{kg/m}^3)}$$
$$= 43.1\,\text{m}^3/\text{d}$$
$$= 43.1\,\text{m}^3/\text{d}(264.172\,\text{gal/m}^3) = 11{,}386\,\text{gpd}$$
$$\text{Waste sludge } (\text{m}^3/\text{d}) = \frac{2200\,\text{kg/d}}{(0.04)(1.02)(1000\,\text{kg/m}^3)}$$
$$= 49.0\,\text{m}^3/\text{d}(12{,}944\,\text{gpd})$$
$$\text{Sludge to digester m}^3/\text{d} = 43.1\,\text{m}^3/\text{d primary} + 49.0\,\text{m}^3/\text{d waste}$$
$$= 92.1\,\text{m}^3/\text{d } (24{,}330\,\text{gpd})$$

c. Compute the volume of the first stage digester.

$$\text{Volume in m}^3 = \frac{3150\,\text{kgVS/d}}{1.28\,\text{kg VS/m}^3 \cdot \text{d}}$$
$$= 2{,}461\,\text{m}^3$$
$$= 2{,}461\,\text{m}^3(35.3147\,\text{ft}^3/\text{m}^3) = 86{,}909\,\text{ft}^3$$

(The volume of the second stage digester also should be 2,461 m^3.)

d. Compute the hydraulic residence time in the primary digester using Eq. (9).

$$\theta = \theta_c = \text{V/Q}$$
$$\theta_c = \frac{2461\,\text{m}^3}{92.1\,\text{m}^3/\text{d}} = 26.7\,\text{d} > 10\,\text{d minimum.}$$

e. Estimate the quantity of digested sludge produced.
From Figure 14.7, $\theta = 27\,\text{d}$, VS = 75%, then volatile solids reduction (VSR) = 54%.

$$\text{Weight of total} = \text{weight of fixed solids (FS)} + \text{weight of volatile solids (VS)}$$
$$= (2{,}200 + 2{,}000)(0.25) + (2{,}200 + 2{,}000)(0.75)(1 - 0.54)$$
$$= 1{,}050\,\text{kg FS/d} + 1{,}449\,\text{kg VS/d}$$
$$= 2{,}499\,\text{kg/d } (5{,}509\ \text{lb/d}) \text{ of digested sludge on a dry weight basis}$$
$$\text{Volume of digested sludge} = \frac{2499\,\text{kg/d}}{(0.08)(1.02)(1000\,\text{kg/m}^3)}$$
$$= 30.6\,\text{m}^3/\text{d } (8{,}090\,\text{gpd})$$

f. Estimate the heating requirements.
Assume a mixed and insulated digester in the northern US and that the temperature of the raw sludge is 10°C.

From Table 14.5 for a unmixed's and noninsulated digester in Northern US:

$$\begin{aligned}\text{Heat loss in digester} &= 4{,}190 \text{ watts}/100\,\text{m}^3 \times 2461\,\text{m}^3 \\ &= 1.03 \times 10^5 \text{ watts} \\ &= 1.03 \times 10^5 \text{ watts } (0.9478\ \text{BTU/sec})(3600\ \text{sec/h}) = 3.51 \times 10^5\,\text{BTU/h}\end{aligned}$$

From Equation (14):

$$H = \text{WC}\,(T_2 - T_1)$$

$$H = \left(\frac{92.1\,\text{m}^3/\text{d} \times 1.02 \times 1000\,\text{kg/m}^3}{24\,\text{h/d}}\right)\left(4200\,\text{J/kg/°C}\right)(35°\text{C} - 10°\text{C})\left(\frac{1\,\text{w}}{3600\,\text{J}}\right)$$

$$= 1.14 \times 10^5\,\text{w}\ (3.89 \times 10^5\ \text{BTU/h})$$

$$\begin{aligned}\text{Total heat requirement} &= 1.03 \times 10^5 + 1.14 \times 10^5 \\ &= 2.17 \times 10^5\,\text{w}\ (7.40 \times 10^5\ \text{BTU/h})\end{aligned}$$

The digester dimensions would be chosen to suit site requirements, standard digester covers, and so on. The heat requirements should be able to be met by burning the digester gas produced. The heat exchanger would be sized to meet the predicted heat requirements.

6.2. Example Using Solids Loading Factor

Estimate the size of the two-stage digesters required to treat the sludge from a community of 80,000 persons. The wastewater flow is assumed to be $30{,}284\,\text{m}^3/\text{d}$ (8.0 MGD). The raw wastewater contains 275 mg/L of total suspended solids (75% volatile) and 250 mg/L of BOD_5. The primary effluent contains 125 mg/L TSS and 165 mg/L BOD_5. Pilot studies have shown that 0.28 kg VSS/kg BOD_5 (0.28 lb VSS/lb BOD_5) excess volatile suspended solids will be generated from the proposed secondary treatment facility. Assume that the thickened raw primary sludge contains about 6% total solids (94% moisture) and that the thickened waste secondary sludge contains 4% total solids (96% moisture). The digested sludge should contain 8% total solids (92% moisture). All sludges are assumed to have a specific gravity of 1.02. Other pertinent design assumptions are as follows:

1. Only the primary digester will be mixed and heated to 35°C.
2. The sludge contains adequate nitrogen and phosphorus for biological growth.

Solution

a. Compute the weight of volatile solids to be added to the digester daily.

$$\text{VS in primary sludge (kg/d)} = \frac{(275\,\text{mg/L} - 125\,\text{mg/L}) \times 30{,}284\,\text{m}^3/\text{d} \times 0.75(1000\,\text{L/m}^3)}{1{,}000{,}000\frac{\text{mg}}{\text{kg}}}$$

$$= 3{,}407\,\text{kg VS/d}\ (7{,}511\,\text{lb VS/d})$$

$$\text{BOD}_5 \text{ to aerator (kg/d)} = \frac{165\text{mg/L} \times 38{,}284\text{m}^3/\text{d}\,(1{,}000\text{L/m}^3)}{1{,}000{,}000\frac{\text{mg}}{\text{kg}}}$$
$$= 4{,}997\,\text{kg/d}\,(11{,}016\,\text{lb/d})$$
$$\text{VSS in waste sludge (kg /d)} = \left(\frac{4997\,\text{kg BOD}_5}{\text{d}}\right)\left(\frac{0.28\,\text{kg VSS}}{\text{kg BOD}_5}\right)$$
$$= \frac{1399\,\text{kgVSS}}{\text{d}}\,(3{,}084\,\text{lb VSS/ d})$$
$$\text{VS to digester (kg/d)} = 3{,}407 + 1{,}399$$
$$= 4{,}806\,\text{kg VS/d}\;\;(10{,}595\,\text{lb VS/d})$$

b. Compute the volume of raw sludge expected.

$$\text{Primary sludge (m}^3\text{/d)} = \frac{3407\,\text{kgVS/d}}{(0.06)(0.75)(1.02)(1000\,\text{kg/m}^3)}$$
$$= 74.2\,\text{m}^3\text{/d}\;\;(19{,}601\,\text{gpd})$$

Assume the waste sludge is 75% volatile.

$$\text{Waste sludge (m}^3\text{/d)} = \frac{1399\,\text{kg VS/d}}{(0.04)(0.75)(1.02)(1000\,\text{kg/m}^3)}$$
$$= 45.7\,\text{m}^3\text{/d}\;\;(12{,}073\,\text{gpd})$$
$$\text{Sludge to digester (m}^3\text{/d)} = 74.2\,\text{m}^3\text{/d primary} + 45.7\,\text{m}^3\text{/d waste}$$
$$= 119.9\,\text{m}^3\text{/d}\;\;(31{,}673\,\text{gpd}).$$
$$\text{TS of combined sludge fed to digester (\%)} = \frac{(0.06)(74.2) + (0.04)(45.7)}{74.2 + 45.7}(100\%)$$
$$= \frac{4.452 + 1.828}{119.9}(100\%)$$
$$= 4.47\%\,\text{TS}$$

c. Compute the volume of the first stage digester.
From Table 14.3, $T = 35°\text{C}$, $\theta_c = 10\,\text{d}$ for design
From Table 14.1, $\theta_c = 10\,\text{d}$, TS = 4.47%, by interpolation
The VS Loading Factor is $3.42\,\text{kg/m}^3\text{/d}$

$$\text{Volume in m}^3 = \frac{4806\,\text{kgVS/d}}{3.42\,\text{kg/m}^3\text{/d}}$$
$$= 1{,}405\,\text{m}^3\;\;(49{,}616\,\text{ft}^3)$$

(The volume of the second stage digester also should be $1{,}405\,\text{m}^3$.)

d. Check the hydraulic residence time using Eq. (9)

$$\theta = \theta_c = V/Q$$

$$\theta_c = \frac{1405\,\text{m}^3}{3.42\,\text{kg/m}^3/\text{d}} = 11.7 \text{ or } 12\,\text{d}$$

e. Estimate the quantity of digested sludge produced.
From Figure 14.8, $\theta = 12$ d, $VS = 75\%$, then volatile solids reduction (VSR) = 57%
Weight of total = weight of fixed + weight of VS

$$\text{Weight of TS in combined raw sludge} = \frac{4806\,\text{kg VS/d}}{0.75}$$

$$= 6{,}408\,\text{kg/d}\,(14{,}127\,\text{lb/d})$$

$$\text{Weight of TS in digested sludge} = (6{,}408)(0.25) + (6{,}408)(0.75)(1 - 0.57)$$

$$= 1{,}602\,\text{kg FS/d} + 2{,}067\,\text{kg VS/d}$$

$$= 3{,}669\,\text{kg/d}(8{,}089\,\text{lb/d}) \text{ of digested sludge on a dry weight basis}$$

$$\text{Volume of digested sludge} = \frac{3669\,\text{kg/d}}{(0.08)(1.02)(1000\,\text{kg/m}^3)}$$

$$= 45\,\text{m}^3/\text{d}\,(11{,}887\,\text{gpd})$$

f. Estimate the quantity of gas produced.
Assume the rate of gas production will be 0.90 m^3 /kg VS destroyed (14.4 ft^3/lb VS destroyed) and that the gas has an energy value of 5850 kg-cal/m^3

$$= (5{,}850\,\text{kg-cal/m}^3)(3.968\,\text{BTU/kg-cal})(\text{m}^3/35.31\,\text{ft}^3)$$

$$= 657\,\text{BTU/ft}^3$$

$$\text{VS destroyed (kg /d)} = (6{,}408\,\text{kgTS/d})(0.75)(0.57)$$

$$= 2{,}739\,\text{kgVS/d}$$

$$\text{Gas produced}(\text{m}^3/\text{d}) = (2{,}739\,\text{kgVS/d})(0.90\,\text{m}^3/\text{kgVS})$$

$$= 2{,}465\,\text{m}^3/\text{d}(87{,}051\,\text{ft}^3/\text{d})$$

$$\text{Energy content of gas produced} = (2{,}465\,\text{m}^3/\text{d})(5{,}850\,\text{kg-cal/m}^3)$$

$$= 1.44 \times 10^7\,\text{kg-cal/d}$$

$$= (1.44 \times 10^7\,\text{kg-cal/d})(3.968\,\text{BTU/kg-cal})$$

$$= 5.71 \times 10^7\,\text{BTU/d}$$

The digester heating requirements would be estimated in the same manner as shown in Section 6.1 above and compared to the energy available from the gas produced.

6.3. *Example Using Modified Anaerobic Contact Process*

Estimate the size of an anaerobic digester required as a first stage biological step in the treatment of the wastes from a meat packing plant. The average raw waste characteristics are as follows:

Flow = 1,500 m^3/d (396,300 gpd)
BOD_5 concentration = 1,200 mg/L
Temperature = 30°C

The digester will be designed as a single stage completely mixed nonrecycle system. The design assumptions are as follows:

a. The digester will be operated at 30°C.
b. θc = 14 d (see Table 14.3).
c. Efficiency of waste use $E_s = 0.80$.
d. The waste contains adequate nitrogen and phosphorus for biological growth.
e. $Y = 0.07$ kg VS/kg BOD_5 used and $k_d = 0.03\ d^{-1}$ at 30°C.

Solution

1. Compute the daily BOD_5 loading.

$$BOD_5 = \frac{1200\,\text{mg/L} \times 1500\,\text{m}^3/\text{d}\,(1000\,\text{L/m}^3)}{1{,}000{,}000\,\text{mg/kg}}$$
$$= 1{,}800\,\text{kg}\,BOD_5/\text{d}\ (3{,}968\,\text{lb/d})$$

2. Compute the digester volume using Equation (9)

$$\theta = \theta_c = V/Q$$
$$\theta_c = 14\,\text{d}$$
$$V = (1{,}500\,\text{m}^3/\text{d})(14\,\text{d})$$
$$= 21{,}000\,\text{m}^3 (741{,}609\,\text{ft}^3)$$

Several digesters in parallel will be needed meet this volume requirement.

3. Computing the volumetric loading:

$$BOD_5\,\text{kg/m}^3/\text{d} = \frac{1800\,\text{kg/d}}{21{,}000\,\text{m}^3}$$
$$= 0.09\,\text{kg/m}^3/\text{d}$$

4. Compute the effluent waste concentration using Equation (4).

$$E_s = \frac{100\,(S_o - S_1)}{S_o}$$
$$S_1 = \frac{100\,(1200) - 80(1200)}{100}$$
$$= 240\,\text{mg/L}$$

5. Computing the quantity of volatile solids produced per day using Equation (8).

$$A = \frac{YQ(S_0 - S_1)}{1 + (k_d)(\theta_c)} \tag{8}$$

$$= \frac{\left(0.07 \frac{\text{kg VS}}{\text{kg BOD}_5}\right)\left(1500 \frac{\text{m}^3}{\text{d}}\right)\left(1000 \frac{\text{L}}{\text{m}^3}\right)\left(1200 \frac{\text{mg}}{\text{L}} - 240 \frac{\text{mg}}{\text{L}}\right)\left(\frac{\text{g}}{1000\,\text{mg}}\right)\left(\frac{\text{kg}}{1000\,\text{g}}\right)}{1 + (0.03\,\text{d}^{-1})(14\,\text{d})}$$

$$= 71\,\text{kg/d}\ (157\,\text{lb/d})$$

6. Using Equation (3) to determine the volume (m^3) of methane gas produced per day and using value obtained from Equation (8) for A.

$$C = 0.35\,(\text{eF-}1.42\text{A})$$

$$= 0.35\,((0.80)(1{,}800) - 1.42(71))$$

$$= 0.35\,(1{,}440 - 100.82)$$

$$= 469\,\text{m}^3\text{/d}\ (16{,}563\,\text{ft}^3\text{/d})$$

7. Estimate the total gas production assuming the gas is 65% methane.

$$\text{Total gas production} = \frac{469}{0.65} = 722\,\text{m}^3\text{/d}\ (25{,}497\,\text{ft}^3\text{/d})$$

7. RECENT DEVELOPMENT IN ANAEROBIC PROCESS

To increase treatment efficiencies, or to shorten treatment period, or to reduce the required size of the treatment unit, developments in technology for anaerobic processes have focused on increasing the density of microorganisms available for treatment. The anaerobic filter (AF), and the expanded bed (EB)/fluidized bed (FB) are examples of this technology.

The AF operates as both suspended growth and attached growth reactor. The AF uses packing, typically plastic media, so that microorganisms can attach to the media and form a biofilm, which in turn increases available biomass for treatment. Plastic media provide large surface area for the microorganism to attach without significantly reducing volume in the reactor. This is illustrated by the specific surface areas of the plastic media ranging from 100 to 187 m^2/m^3 and the void volumes for the plastic media ranging from 85% to 95% (23–26). Additionally, the media aid in maintaining suspended solids in the treatment unit. The plastic media occupy between 33% and 70% of the volume of the treatment unit with the remaining volume for biogas storage (25, 27). Recycling of the effluent is typically used to maintain a uniform hydraulic loading to the treatment unit. The flow through for an AF can either be upflow or downflow. Anaerobic filters operated in the upflow pattern, generally has more biomass in suspension than AFs operated in the downflow pattern. With either flow pattern, gas is collected in the top of the reactor.

During startup, Puňal et al. (23) determined that limiting nitrogen concentration in the influent during the first two weeks followed by a nitrogen balance influent promotes bacterial biofilm formation. Show and Tay (24) determined that the performance of AF is effective by the surface texture or porosity of the media at organic loading rate greater than 4 g COD/L-d

(4 kg COD/m^3-d). It was observed that media with surface texture performed significantly better than media with smooth surface. It was also observed that media with higher porosity performed better than media with lower porosity. At organic loading rate of 8 g COD/L-d (8 kg of COD/m^3-d) the media with either surface texture or high porosity achieve higher than 90% COD removal, whereas the smooth surface media or the lower porosity media achieved only approximately 75% COD removal. During startup of AF, Smith et al. (25) showed that the upflow velocity should be in the range of 8 to 10 meter per day (m/d) to maintain biomass below the media, encourage gas production and reduce the loss of solids in the effluent. As the concentration of biomass increases the upflow velocity can be increased up to 17 m/d, which will reduce the formation of dead zones by agitating the sludge bed.

The anaerobic EB reactor and FB reactor operate as attached growth reactor. The microorganisms are supported on inert media, which have large amount of surface area. These media typically have low void volumes; occupy more volume of the reactor, which results in the reactor providing shorter hydraulic detention time than reactor filled media with larger void volume. This is illustrated by the specific surface area for the media ranging from 4,000 to 10,000 m^2/m^3, but having void volumes ranging from 45% to 55% (26). Silicon sand or granular activated carbon are commonly used as media (26, 28). The flow pattern for anaerobic expanded bed reactor is upwards at velocities that are sufficient to expand the bed of media by 10% to 30%, whereas in a fluidized bed reactor is also upward reactor but with a velocity high enough to expand the media from 25% to 300%. In the EB reactor, the media with bio-film are partly supported by the fluid and partly supported by the adjacent media. The higher velocities, used in the FB reactor, allow the media with the bio-film to be fully supported by fluid. As the media accumulate biomass, their density decreases. Along with upward flow of the fluid these lighter media with biomass will rise to the top of the fluid. Utilizing this phenomenon solids wasting is generally performed from the top of the reactor by removing these lighter bio-media. The biomasses are shear off the media and the media are returned to the reactor.

Maloney et al. (28) operated a commercial-scale AFB reactor filled with granular activated carbon (GAC). The unit was 0.5 meter in diameter and 4.6 meter tall. The AFB was operated at steady-state flows of 0.03 to 0.05 liters per sec, which provide averaged hydraulic retention time of 10 hours. These flow rates provide upflow velocities of 13 to 22 m/d. The average organic concentrations in the influent were 142 mg/L of dinitrotolune (DNT), 377 mg/L of ether, 2,410 mg/L of alcohol and 9,200 mg/L of COD. The results of this study showed that the AFB with GAC can lower the concentration of difficult to degrade organic dinitrotolune (DNT) to less than 0.08 mg/L or greater than 99% removal.

A suspended growth process that has been developed is the upflow anaerobic sludge blanket (UASB) (12, 26, 27, 29–31). The process incorporates a bottom feed reactor that distributes the flow uniformly across the cross-section of the reactor. As the flow travels upwards through the reactor, a blanket, the dense slurry of granular biomass is formed. Treatment occurs as wastewater passes through the blanket and the granular biomass break down large organic molecules to water, carbon dioxide and methane, which also include intermediate steps. The type wastewater being treated can cause the makeup of the blanket to vary. Wastewater containing suspended solids or other matter that are not being by trapped by the granular biomass will pass through and form a less dense blanket above denser granular biomass

blanket or be discharge with effluent. This less dense blanket is referred to as the flocculent sludge blanket.

The granular blanket is suspended in the lower section of the reactor by the upward flow through the reactor. The upper portion of the reactor contains the flocculent blanket (if it forms), a zone for settling of solids and gas/solid separating, which allows solids to be returned to the blanket and solids/gas particles to be degasified. After solids separation, effluent and the biogases are removed. As the treatment process continues, the biomass continues to grow; solids inventory increases, which result in the increase in the depth of the blanket. As depth of the blanket increases due to solids buildup, the effectiveness of the separator will decrease. This buildup can interfere with solid and gas separation and effluent quality. To reduce the blanket or reduce the solids buildup in the process, solids are wasted. Solids are withdrawal at blanket level. The depth of the blanket varies as biomass is increasing, so multiple ports must be provided at various depths.

UASB is also considered to have good mixing characteristics without utilizing internal mechanical mixing devices. The mixing occurs in the sludge blanket and is a result of a combination of the influent distribution and gassing that results from anaerobic digestion process. The mixing also aids in the formation and the maintenance of the granular biomass.

Yu, Tay and Fang (31) reported that optimum granulation occurred when the calcium concentration ranged from 150 to 300 mg/L and the COD influent concentration is 4,000 mg/L. At lower calcium concentrations minimal granulation occurred and at higher concentration there was a tendency of cementation of sludge blanket. The optimal granulation is also dependent on the influent COD concentration, since it has effect on the calcium carbonate precipitate in the granules. In this study, the sizes of the granules ranged from 0.2 to 0.6 mm after 30 days and the range reached 1 to 2 mm after 60 days of operation. It was also determined in this study, the calcium concentration in granules was proportional to the influent calcium concentration and calcium carbonate was the main calcium precipitate in granules. The authors concluded that specific activity of granules decreased with increasing calcium concentration in influent. It was noted by the authors that higher calcium concentration led to larger granules with higher ash content, which reduced mass transfer. The addition of low concentrations of calcium into the influent improves the formation of granules in an UASB by enhancing adsorption, adhesion and multiplication. A modification to UASB reactor is expanded granular sludge bed (EGSB), which combines the ultra high loading of fluidized process and granulation of biomass in the upflow anaerobic sludge blanket process (26, 32).

Zoutberg and de Been (32) reports on the full-scale EGSB installation that treated formaldehyde operated at superficial upflow velocity in the reactor of 9.4 m/h, which is much high than conventional UASB reactor maximum superficial upflow velocity of 1 m/h. This higher velocity in the EGSB allows for high recirculation flow rate and low influent flow rate, which diluted the raw wastewater (formaldehyde) 30 times with anaerobic effluent. The volumetric total COD loading to the EGSB was 17 kg COD/m^3-d. The influent COD concentration ranged from approximately 5,000 to 45,000 mg/L and effluent ranged from 350 to 900 mg/L. This EGSB achieved 98% removal. Typical performance data for anaerobic processes are presented in the Table 14.6. These values are approximate and are intended to give a range that is applicable for the process.

Table 14.6
Performance data for anaerobic processes

Process	Hydraulic detention time (h)	Organic loading (kg COD/m^3/day)	Removal COD (%)
Anaerobic filter	10–20	2–8	60–90
Expanded bed	5–10	5–10	70–95
Upflow anaerobic sludge blanket	12–120	5–15	60–90
Expanded granular sludge bed	4–72	5–30	70–95

To improve operations of anaerobic treatment process, developments in technology for anaerobic processes have focused on the physical shapes or attributes of the reactor. The egg-shape digester (ESD) and the waffle floor design are examples of this technology (21, 33, 34).

A development in the United States is the use of the steep-sided conical bottom tank with converging top cone design for anaerobic digesters, which have been used extensively in Europe since the 1950s. The EDS is an example this type of digester. In mid 1970, the first egg-shape digester was built in the United States and is becoming more common, because the bottom shape eliminates the need for cleaning.

Early ESD designs were constructed from reinforced concrete with 37° conical side slopes, whereas reported recent designs have been constructed from steel with 37° to 45° conical side slopes (34). Both designs provide small bottom area and the step sloped sides that concentrates the settled grit to a small area and this in turn provides a central location to effectively remove the grit. As a result, grit does not accumulate in the digester and the effective treatment volume is not reduced, whereas in a mild sloped designed bottom digester, which results in a larger bottom area, the settled grit is spread over a larger area, making it more difficult to remove the settled grit. Therefore, as grit accumulating in the digester, the treatment volume in digester is reduced. Another benefit of the ESD, is that the shape minimizes the liquid surface area. This small top area limits the tendency for scum buildup and debris accumulation. Scum that accumulates at the liquid surface can be kept fluid with a mixer and can be easily removed.

Another approach to improve grit removal, to eliminate grit accumulation and maintain process volume, is to provide a waffle bottom design (a bottom with multiple sections) on circular digester with multiple grit removal drains (21, 34). This bottom design involves a center cones section with 45° conical side with a center drain to remove settled grit. The remaining portion of the digester bottom is divided into section from the outer edge of the center cone to the outer edge of the digester bottom. These multiple outer sections are provided with steep side slopes and bottoms that are sloped up to 20° to the outer rim, where grit removal drains are provided to remove settled grit.

Temperature-phased anaerobic digestion (TPAD) (35–40) has been developed to meet the requirements for pathogen reduction (density of fecal coliform in the biosolids must be less than 1,000 most probable number per gram of total solids) and vector attraction reduction (38% reduction in volatile solids content of the biosolids) as required in Title 40 of the Code of Federal Regulations (CFR), Part 503 (42). The anaerobic digester in the first stage

is operated at temperatures ranging from 50° to 60°C (thermophilic) and the second stage anaerobic digester is operated at temperature range of 34° to 38°C (mesophilic). In selecting operating temperatures and SRTs for the stages, a balance must be obtained between the pathogen reduction, heat exchanger size, and energy consumption. As reported by Han and Dague (38), the fecal coliform destruction is a primary function of the higher temperature in the thermophilic phase. Their bench-scale study showed that the thermophilic phase (55°C) achieved 99.9998% reduction in fecal coliforms when the SRTs varied from 3.3 to 5 days with a constant operating temperature of 55°C. Varying SRTs for the mesophilic phase (35°C) from 6.7 to 10 days, the researchers observed that TPAD process achieved 39% VS destruction for TPAD process at SRT of 6.7 days and 53.2% VS destruction at SRT of 10 days. The research also reported that a single stage mesophilic process (35°C) with SRT of 10 days achieved 32% VS destruction and with SRT of 15 days achieved 46.8% volatile destruction, but it did not achieve the fecal coliform reduction.

Operating parameters for thermophilic/mesophilic TPAD process like all other biological processes are dependent on the wastewater characteristics both hydraulic and organic loadings and the margin of safety required by operating personnel. The design engineer must evaluate these loadings and provide the operating personnel sufficient flexibilities in the process so that the treatment goals can be achieved. Table 14.7 presents reported parameters for the thermophilic/mesophilic TPAD process.

To improve treatment efficiency and operations of anaerobic treatment process, researchers have examined the addition of media into anaerobic unit (43), and developed control system to regulate the COD (44), respectively. Mathematical model has also been developed for predicting the conversion of complex organic into biogas in a batch reactor (41).

Taricska (43) demonstrated that a continuous flow, two-stage anaerobic/aerobic treatment process can effectively treat synthetic milk wastewater which had total organic carbon (TOC) concentrations of approximately 900, 2,300 and 3,800 mg/L. The anaerobic unit with media had hydraulic detention time of 6.5 days and the aerobic units with media had hydraulic detention times of 5, 10 and 15 days. Media addition improved the TOC removal efficiency of the anaerobic reactor as much as 16.4%. The author showed how these processes could be adapted to anaerobic/aerobic lagoon system with media addition to anaerobic lagoon. To aid engineers, design curves were developed. Gray et al. (49) also present additional application of anaerobic digestion of food waste.

Mendez-Acosta et al. (44) proposed a control system for anaerobic digestion. The proposed system is a dynamic output feedback control for the regulation of the chemical oxygen demand (COD) in anaerobic digestion process. The control law has two different structures: (a) nonlinear if a nominal value of the influent COD concentration is available; and (b) linear if such a nominal value is not available. To achieve the COD regulation, the control law includes high gain observer for dynamic estimator, which can induce undesired effects such as the so-called peaking phenomena. This phenomenon produces large overshoots on the control inputs and leads to windup behavior as a result of constraints on the control input. The authors showed that two schemes can improve the performance of the control system. The two schemes are: (a) an antiwindup scheme to consider the constraints in the control input; and (b) a fuzzy-based gain scheduling to tune the control parameter.

Table 14.7
Performance parameter for TPAD process

Plant	Author	Type of sludge	Thermophilic SRT (d)	Mesophilic SRT (d)	VS Destruction (%)	Density fecal coliform (MPN/g TS)
Cologne Germany	Schafer & Farrell (37)	WAS	7	27	43	
Wilhelmshaven Germany	Schafer & Farrell (37)	Primary & WAS	3–5	13–15	54	
Belmont WWTP, Indiana	Shimp et al. (35)	Primary & WAS	2	10	>38	<1,000
	Chao (39)	Pulp & Paper	2.5 to 5	7.5 to 15	38.2	
	Vanderburgh and Ellis (40)	Primary & WAS			>38	<1,000
	Oles et al. (41)	Municipal	2 to 3	12 to 15		

Researchers (41) developed a structural mathematical model of anaerobic conversion of complex organic materials in nonideally cyclic-batch reactors for biogas production. The model was applied to anaerobic digestion of cattle manure and showed good correlation to experimental data. More recent developments can be found from the literature (48, 49).

NOMENCLATURE

A = VS produced (kg/d)
A = Surface area of tank element, (m^2 (ft^2))
A = Excess microorganism production rate (kg/d (lb/d))
C = Mean specific heat of raw sludge (4200 J/kg/°C (1 BTU/lb/°F))
C = Methane produced (m^3/d (ft^3/d))
E_S = Process efficiency (%)
e = Efficiency of waste use (%)
F = BOD_5 added (kg/d (lb/d))
H = Amount of heat required (J (BTU))
k = Maximum substrate removal rate (d^{-1})
k_d = Microorganisms decay coefficient (d^{-1})
K_S = Waste concentration at which the rate of waste use per unit weight of microorganism is one half the maximum rate (mg/L)
Q = Heat loss from the tank (w (BTU/h))
Q = Waste flow rate (m^3/d (gpd))
Q_r = Return sludge flow rate (m^3/d (gpd))
U = Food to microorganism ratio
r = Return sludge ratio (Q_r/Q)
S_0 = Influent substrate concentration (mg/L)
S_1 = Effluent substrate concentration (mg/L)
T_1 = Temperature of raw sludge entering tank or temperature outside the tank (°C (°F))
T_2 = Temperature of sludge in the tank or temperature inside the tank (°C (°F))
U = Heat transfer coefficient (w/m^2/°C (BTU/h/ft^2/°F))
V = Volume of the reactor (m^3 (ft^3))
X = Mass of volatile solids in reactor (kg (lb))
X_r = Mass of volatile solids in the return sludge (kg (lb))
Y = Cell yield (mg/mg (lb/lb))
W = Weight of sludge entering the tank per hour (kg (lb))
θ = Hydraulic retention time (d)
θ_c = Mean cell residence time (d)
θ_c^m = Minimum mean cell residence time (d)

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15
Aerobic Digestion

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CONTENTS

Abstract Aerobic digestion is the biochemical oxidative stabilization of wastewater sludge in open or closed tanks that are separate from the liquid process system. This method of digestion is capable of handling waste activated, trickling filter, or primary sludges as well as mixtures of the same. Aerobic digestion is based upon endogenous respiration, where in the absence of suitable substrate food, microorganisms begin to digest their own protoplasm to obtain energy. Cell tissue is aerobically oxidized to carbon dioxide, water, and ammonia or nitrates. Some of the energy released by the microbial degradation is used to form new cellular material, but the majority is released as heat; thus the aerobic oxidation process is exothermic. In a large facility, it may be feasible or desirable to digest primary sludge anaerobically, and secondary sludge aerobically. Following process description, the chapter covers process variations, design considerations, process performance, costs and worked out design examples.

Key Words Aerobic digestion • endogenous respiration • exothermic • parameters • performance • design • costs.

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1. INTRODUCTION

Both aerobic and anaerobic digestion processes are being used in new designs for treating biological sludges; there are advantages and disadvantages to both systems. Before a specific choice can be made, waste characteristics, general climatic conditions, type of sludge handling equipment, and the capacity of the facility must be considered. In a large facility, it may be feasible or desirable to digest primary sludge anaerobically, and secondary sludge aerobically.

Aerobic digestion is the biochemical oxidative stabilization of wastewater sludge in open or closed tanks that are separate from the liquid process system. This method of digestion is capable of handling waste activated, trickling filter, or primary sludges as well as mixtures of the same. The aerobic digester operates on the same principles as the activated sludge process. As food, is depleted, the microbes enter the endogenous phase and the cell tissue is aerobically oxidized to CO_2, H_2O, NH_4^+, NO_2^-, and NO_3^- (1).

Air or oxygen can be supplied by surface aerators or by diffusers (2). Other equipment may include sludge recirculation pumps and piping, mixers and scum collection baffles (3). Aerobic digesters are designed similarly to rectangular aeration tanks and use conventional aeration systems, or employ circular tanks and use an eductor tube for deep tank aeration.

Studies on aerobic digestion of municipal wastewater sludge have been conducted since the early 1950's (4, 5). Early studies (6, 7) indicated that aerobic digestion performed as well as, if not better than, anaerobic digestion in reducing volatile solids in sludge. Aerobic digestion processes were economical to construct, had fewer operating problems than anaerobic processes, and produced a digested sludge that drained well. By 1963, at least one major equipment supplier (8) had approximately 130 installations in plants with flow from 10,000 to 100,000 gpd (37.8 to 378 m^3/d). By the late 1960's and early 1970's, consulting engineers across the country were specifying aerobic digestion facilities for many of the plants they were designing. As of early 1980's, numerous plants used aerobic digestion, and several of them are quite large.

2. PROCESS DESCRIPTION

2.1. Microbiology

Aerobic digestion of municipal wastewater sludges is based on the principle that, when there is inadequate external substrate available, microorganisms metabolize their own celluar mass. In actual operation, aerobic digestion involves the direct oxidation of any biodegradable matter and the oxidation of microbial cellular material by organisms. These two steps are illustrated by the following reactions (9):

$$\text{Organic matter} + O_2 \rightarrow \text{Cellular material} + CO_2 + H_2O \quad (1)$$

$$\text{Cellular material} + O_2 \rightarrow \text{Digested sludge} + CO_2 + H_2O \quad (2)$$

The process described by Equation (2) is referred to as endogenous respiration, which is normally the predominant reaction in aerobic digestion.

2.2. Advantages

Various advantages have been claimed for aerobic digestion over other stabilization techniques, particularly anaerobic digestion (10). Based on all current knowledge, the following advantages can be cited for properly designed and operated aerobic digestion processes (11–15):

1. Have capital costs generally lower than for anaerobic systems for plants under 5 MGD (220 L/s)
2. Are relatively easy to operate compared to anaerobic systems
3. VSS is reduced to 40–50 percent, nearly equivalent to that for anaerobic
4. Do not generate nuisance odors
5. Will produce a supernatant low in BOD_5, suspended solids, and ammonia nitrogen
6. Reduce the quantity of grease in the sludge mass
7. A relatively stable humus like end product is produced
8. Reduce the number of pathogens to a low level under normal design. Under auto-heated thermophilic design, many systems provide 100 percent pathogen destruction.

2.3. Disadvantages

As with any process, there are also certain disadvantages. In aerobic digestion processes, the disadvantages are:

1. Usually produce a digested sludge with very poor mechanical dewatering characteristics
2. Have high power costs to supply oxygen, even for very small plants
3. Are significantly influenced in performance by temperature, location, and type of tank material
4. No heavy metal removal
5. Lack of useful by-product (no methane).

3. PROCESS VARIATIONS

Both one and two tank systems are used. Small plants often use a one tank batch system with a complete mix cycle followed by settling and decanting (to help thicken the sludge). Larger plants may consider a separate sedimentation tank to allow continuous flow and facilitate decanting and thickening. Air may be replaced with oxygen.

3.1. Conventional Semi-Batch Operation

Originally, aerobic digestion was designed as a semi-batch process, and this concept is still functional at many facilities. Solids are pumped directly from the clarifiers into the aerobic digester. The time required for filling the digester depends on available tank volume, volume of waste sludge, precipitation, and evaporation. During the filling operation, sludge undergoing digestion is continually aerated. When the tank is full, aeration continues for two to three weeks to assure that the solids are thoroughly stabilized. Aeration is then discontinued and the stabilized solids settled. Clarified liquid is decanted, and the thickened solids are removed at a concentration of between two and four percent. When a sufficient amount of stabilized sludge and/or supernatant have been removed, the cycle is repeated. Between cycles, it is customary to leave some stabilized sludge in the aerator to provide the necessary microbial population

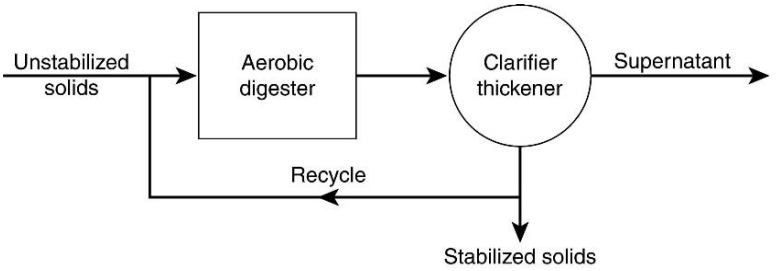

Fig. 15.1. Flow diagram for a conventional aerobic digestion process (*Source*: US EPA) (9).

for degrading the wastewater solids. The aeration device need not operate for several days, provided no raw sludge is added.

Many engineers have tried to make the semi-batch process more continuous by installing stilling wells to act as clarifiers in part of the digester. This has not proven effective (15–17).

3.2. Conventional Continuous Operation

The conventional continuous aerobic digestion process closely resembles the activated sludge process as shown on Figure 15.1. As in the semi-batch process, solids are pumped directly from clarifiers into the aerobic digester. The aerator operates at a fixed level, with the overflow going to a solids-liquid separator. Thickened and stabilized solids are either recycled back to the digestion tank or removed for further processing (9).

The process is highly reliable. It is less sensitive to environmental factors than anaerobic digestion and requires less laboratory control and daily maintenance. It is relatively resistant to variations in loading, pH and metals interference. Lower temperatures require much longer detention times to achieve a fixed level of VSS reduction. However, performance loss does not necessarily cause an odorous product (18). Maintenance of the DO at 1 to 2 mg/L with adequate detention results in a sludge that is often easier to dewater (except on vacuum filters) (19, 20).

3.3. Autothermal Thermophilic Aerobic Digestion (Using Air)

A new concept that is receiving considerable attention in the United States is the auto-heated aerobic digestion process (13, 21). This autothermal thermophilic aerobic digestion using air is a form of aerobic digestion that operates in the thermophilic temperature range (greater than 45°C) using air as the source of oxygen to aerate the sludge. The operation is autothermal; that is, the heat required for the increase in temperature is supplied completely from the exothermic breakdown of organic and cellular material occurring during aerobic digestion. The increased temperature, in turn, reduces the required retention time for a given amount of solids reduction (22–27). The digester tanks are covered and insulated to minimize heat losses from the system.

In this process, sludge from the clarifiers is usually thickened to provide a digester feed solids concentration of greater than four percent. The heat liberated in the biological

degradation of the organic solids is sufficient to raise the liquid temperature in the digester to as high as 140°F (60°C) (13).

The high temperatures reached in the digester may result in virtually complete destruction of pathogens and eliminate the need for further disinfection (28, 29). Thermophilic conditions can be reached in most climates and will require a much shorter retention time than unheated aerobic digestion or anaerobic digestion. At temperatures above 50°C, a high degree of digestion and of solids removal can be achieved with less than 8 days' retention. The high temperatures also decrease oxygen requirements because of the inhibition of nitrification. In general, aerobic digestion produces a supernatant with lower organic loadings than anaerobic digestion. The process may improve the settleability and dewatering characteristics of sludge. The simplicity of operation may be suitable for use by small treatment plants. The process may have application in cold climates where conventional aerobic digestion is ineffective or requires excessively long detention times. The commonly used design temperature is in the range of 45 to 70°C and the recommended retention time is from 2 to 10 days (30).

Advantages claimed for this mode of operation are (13, 21, 30–32):

1. Higher rates of organic solids destruction
2. Smaller volume requirements
3. Production of a pasteurized sludge
4. Destruction of all weed seeds
5. 30 to 40 percent less oxygen requirement
6. Improved solids-liquid separation through decreased liquid viscosity.

Disadvantages cited for this process are that:

1. It must incorporate a thickening operation
2. Mixing requirements are higher because of the higher solids content
3. Non-oxygen aerated systems require extremely efficient aeration and insulated tanks.

For further detailed discussion on temperature effect, metabolic inhibitors and design issues of thermophilic digestion, the reader is directed to References (33–36).

3.4. Autothermal Thermophilic Aerobic Digestion (Using Oxygen)

Autothermal thermophilic oxygen digestion using oxygen is a form of aerobic digestion that operates in the thermophilic (more than 45°C) temperature range and utilizes pure oxygen instead of air to aerate the sludge. The operation is autothermal; that is, the heat required for the increase in temperature is supplied completely from the exothermic breakdown of organic and cellular material occurring during aerobic digestion. The increased temperatures, in turn, reduce the required retention times in the digesters to achieve a given amount of solids reduction. The digester tanks are covered to minimize heat losses from the system. Heat losses are also reduced in pure oxygen systems because there is little exhaust gas to remove the heat generated by the process (37). The equipment for pure oxygen thermophilic aerobic digestion is similar to that of aerobic digestion with the addition of digester covers and an oxygen generator.

This process may have greatest applications where pure oxygen activated sludge processes are used. The high temperatures used by the process may result in virtually complete destruction of pathogens, and eliminate the need for further disinfection. In colder climates the

process will have much shorter retention times than other digestion processes. At temperatures above 45°C a high degree of digestion can be obtained with less than five days retention. The high temperatures decrease oxygen requirements because of the inhibition of nitrification. In general, aerobic digestion produces a supernatant with lower organic loadings than anaerobic digestion. The danger of methane explosions is also reduced.

The process may not be applicable to conventional unthickened waste activated sludges because of the large amount of heat required to raise WAS (waste activated sludge) at 0.5 percent solids to thermophilic temperatures. The process has high operating costs (primarily to supply oxygen). No useful byproducts such as methane are produced. Oxygen aerobic digestion in the mesophilic temperature range does not appear to be cost effective, but in the thermophilic range the reduced O_2 requirements and smaller reactor volume may enable the process to be competitive with other forms of digestion, particularly when a pathogen-free sludge is desired.

Single or two stage systems could be used with a retention time of five days or less and at operating temperatures in the thermophilic range of 45 to 60°C.

4. DESIGN CONSIDERATIONS

Factors to be considered during the design process are characteristics of the sludge, residence time, solids loading criteria, energy requirement for mixing, environmental conditions, and process operation.

4.1. Temperature

Since the majority of aerobic digesters are open tanks, digester liquid temperatures are dependent on weather conditions and can fluctuate extensively. As with all biological systems, lower temperatures retard the process while higher temperatures speed it up. A large number of studies on aerobic digestion of municipal sludges as a function of liquid temperature have been carried out (38–50). When considering temperature effects in system design, one should design a system to minimize heat losses by using concrete instead of steel tanks, placing the tanks below rather than above grade, and using sub-surface instead of surface aeration. Design should allow for the necessary degree of sludge stabilization at the lowest expected liquid operating temperature, and should meet maximum oxygen requirements at the maximum expected liquid operating temperature.

4.2. Solids Reduction

A major objective of aerobic digestion is to reduce the mass of solids for disposal. This reduction is assumed to take place only with the biodegradable content of the sludge, though some studies (51, 52) have shown that there may be destruction of the non-organics as well. In this discussion, solids reduction will pertain only to the biodegradable content of the sludge.

The change in biodegradable volatile solids can be represented by a first order biochemical reaction:

$$\frac{\mathrm{d}M}{\mathrm{d}t} = -K_{\mathrm{d}}M \tag{3}$$

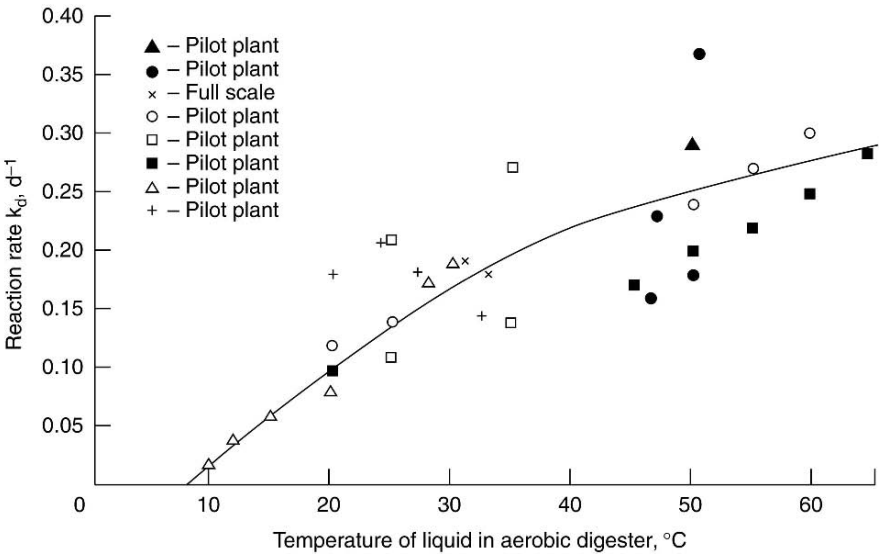

Fig. 15.2. Aerobic digestion reaction rate (K_d) as a function of sludge temperature (*Source:* US EPA) (9).

where

dM/dt = rate of change of biodegradable volatile solids per unit of time, mg/L/d
K_d = reaction rate constant, d^{-1}
M = concentration of biodegradable volatile solids remaining at time t in the aerobic digester, mg/L
t = time, d

The time t in Equation (3) is actually the sludge age or solids residence time in the aerobic digester. Depending on how the aerobic digester is being operated, time t can be equal to or considerably greater than the theoretical hydraulic residence time. The reaction rate term K_d is a function of sludge type, temperature, and solids concentration. It is a pseudo-constant, since the term's value is the average result of many influences (9). Figure 15.2 shows a plot of various reported K_d values as a function of the digestion temperature. The data shown are for several different types of waste sludge, which partially explains the scatter. Furthermore, there has been no adjustment in the value of K_d for sludge age. The line drawn through the data points represents an overall average K_d value.

Little research has been conducted on the effect of solids concentration on reaction rate K_d. The results of one study (49) with waste activated sludge at a temperature of 68°F (20°C) are shown on Figure 15.3, which indicates that K_d decreases with increasing solids concentration. Stabilization is not complete until there has been an extended period of primarily endogenous respiration (typically 15 to 20 days). Up to 80 percent of the cell tissue may be oxidized; the remaining fractions contain inert and nonbiodegradable materials (9).

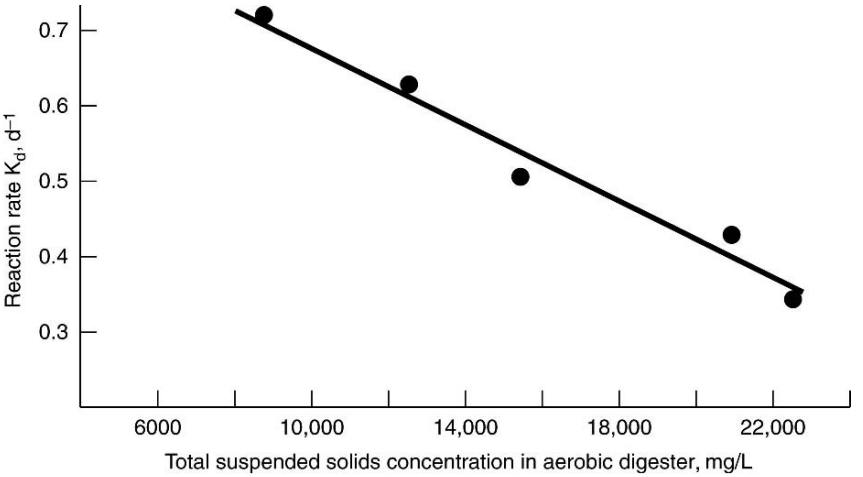

Fig. 15.3. Aerobic digestion reaction rate (K_d) as a function of solids concentration (*Source:* US EPA) (9).

4.3. Oxygen Requirements

Activated sludge biomass is most often represented by the empirical equation $C_5H_7NO_2$. The aerobic digestion of the sludge is usually depicted as follows:

$$C_5H_7O_2N + 5\,O_2 \rightarrow 5\,CO_2 + 2\,H_2O + NH_3 \tag{4a}$$

Under the prolonged periods of aeration typical of the aerobic digestion process, the NH_3 in the presence of excess oxygen is further oxidized to NO_2^- to NO_3^- as shown in Equation (4b):

$$C_5H_7NO_2 + 7\,O_2 \rightarrow 5\,CO_2 + 3\,H_2O + H^+ + NO_3^- \tag{4b}$$

Hypothetically, this equation indicates that 1.98 pounds (0.898 kg) of oxygen are required to oxidize one pound (0.45 kg) of cell mass. From pilot and full-scale studies, however, the pounds of oxygen required to degrade a pound of volatile solids were found to be 1.74 to 2.07 (0.789 to 0.939 kg). For mesophilic systems, a design value of 2.0 is recommended. For autothermal systems, which have temperatures above 113°F (45°C), nitrification does not occur and a value of 1.45 is recommended (13, 21, 31).

The actual specific oxygen utilization rate, pounds oxygen per 1,000 pounds volatile solids per hour, is a function of total sludge age and liquid temperature (14, 47, 51). In one study, Ahlberg and Boyko (14) visited several operating installations and developed the relationship shown on Figure 15.4. Specific oxygen utilization is seen to decrease with increase in sludge age and decrease in digestion temperature (9).

Field studies have also indicated that a minimum value of 1.0 mg of oxygen per liter should be maintained in the digester at all times (14).

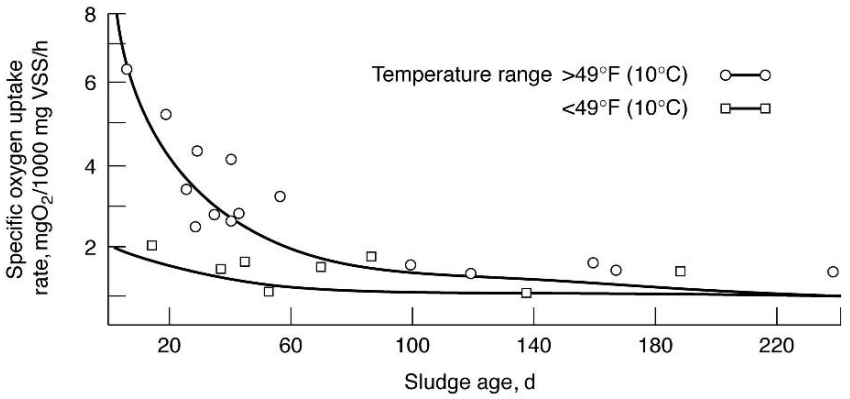

Fig. 15.4. Effect of sludge age and temperature on oxygen uptake rates (*Source:* US EPA) (9).

4.4. Mixing

Mixing is required in an aerobic digester to keep solids in suspension and to bring deoxygenated liquid continuously to the aeration device. Whichever of these two requirements needs the most mixing energy controls the design.

According to treatment plants experience, levels ranging from 0.5 to 4.0 horsepower per 1,000 ft^3 of tank volume (13 to 106 kw/1,000 m^3) are satisfactory. Designers should consult an experienced aeration equipment manufacturer for assistance in design.

4.5. pH Reduction

The effect of increasing detention time on pH of sludge in the aerobic digester during mesophilic temperature range operation is shown on Figure 15.5.

The drop in pH and alkalinity is caused by acid formation that occurs during nitrification (9). Although at one time the low pH was considered inhibitory to the process, it has been shown that the system will acclimate and perform just as well at the lower pH values (42, 47, 53, 54). It should be noted that if nitrification does not take place, pH will drop little if at all. This could happen at low liquid temperatures and short sludge ages or in thermophilic operation (21). Nitrifying bacteria are sensitive to heat and do not survive in temperatures over 113°F (45°C) (55).

4.6. Dewatering

Although there are published reports of excellent operating systems (48, 56) much of the literature on full-scale operations has indicated that mechanical dewatering of aerobically digested sludge is very difficult (39, 44, 57). Furthermore, in most recent investigations, it is agreed that the dewatering properties of aerobically digested sludge deteriorate with increasing sludge age (38, 44, 58, 59). Unless pilot plant data indicate otherwise, it is recommended that conservative criteria be used for designing mechanical sludge dewatering facilities for

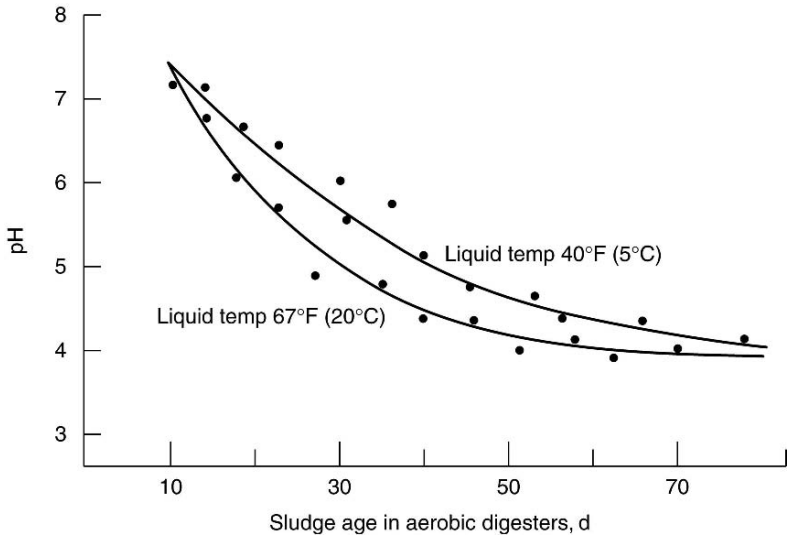

Fig. 15.5. Influence of sludge age on pH during aerobic digestion (*Source:* US EPA) (9).

aerobically digested sludge. As an example, a designer would probably consider designing a rotary vacuum filter for a production rate of 1.5 lb of dry solids/ft^2/h (7.4kg/m^2/h), a cake solids concentration of 16 percent, with a $FeC1_3$ dose of 140 lb (63.5 kg), and a lime dose (CaO) of 240 lb (109 kg). This assumes an aerobic solids concentration of 2.5 percent solids.

5. PROCESS PERFORMANCE

5.1. Total Volatile Solids Reduction

Solids destruction has been shown to be primarily a direct function of both basin liquid temperature and the length of time during which the sludge was in the digester (54, 60, 61). Figure 15.6 is a plot of volatile solids reduction versus the parameter degree-days. Data were taken from both pilot and full-scale studies on several types of municipal wastewater sludges (9). Figure 15.6 indicates that, for these sludges, volatile solids reductions of 40 to 50 percent are obtainable under normal aeration conditions.

Up to 85 percent reduction in pathogens could be attained in mesophilic digestion while under thermophilic conditions it is possible to get complete removal of all pathogens.

5.2. Supernatant Quality

The supernatant from aerobic digesters is normally returned to the head end of the treatment plant. Typical supernatant quality is as follows:

(a) Suspended solids 100 to 12,000 mg/L
(b) BOD_5 50 to 1700 mg/L
(c) Soluble BOD_5 4 to 200 mg/L
(d) COD 200 to 8000 mg/L
(e) Kjeldahl N (TKN) 10 to 400 mg/L

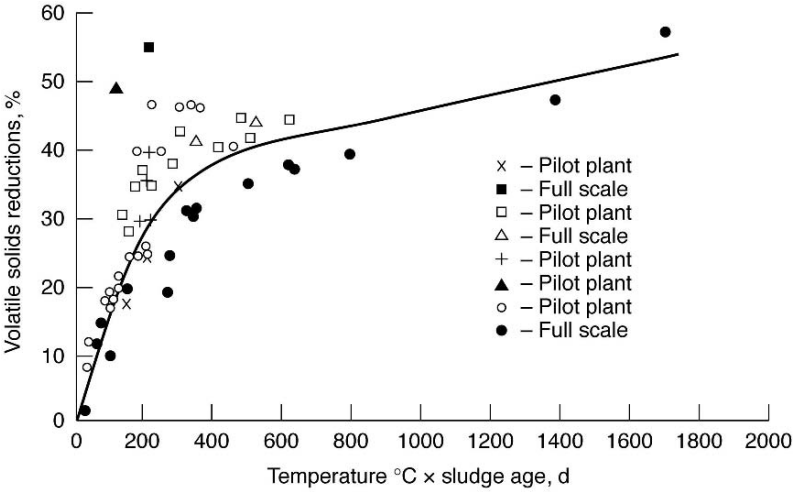

Fig. 15.6. Volatile solids reduction as a function of sludge temperature and age (*Source:* US EPA) (9).

(f) Total P 20 to 250 mg/L
(g) Soluble P 2 to 60 mg/L
(h) pH 5.5 to 7.7

For further details on aerobic digestion, especially when dealing with mixed sludges, the presence of chemical oxidizing agents, solid phase aerobic process and process modeling to simulate optimal operating conditions, the reader is referred to References (34, 54, 62–64).

6. PROCESS DESIGN

Design criteria for aerobic digestion commonly include the following parameters. Solids retention time (SRT) required for 40% VSS reduction: 18 to 20 days at 20°C for mixed sludges from activated sludge or trickling filter plant, 10 to 16 days for waste activated sludge only, 16 to 18 days average for activated sludge from plants without primary settling; volume allowance: 3 to 4 ft^3/capita; VSS loading: 0.02 to 0.4 lb/ft^3/d; air requirements: 20 to 60 ft^3/min/1000 ft^3; minimum DO: 1 to 2 mg/L; energy for mechanical mixing: 0.75 to 1.25 hp/1000 ft^3; oxygen requirements: 2 lb/lb of cell tissue destroyed (includes nitrification demand), 1.6 to 1.9 lb/lb of BOD removed in primary sludge (14, 15, 25, 36, 65–68).

6.1. Input Data

(a) Primary sludge production, lb/d
(b) Secondary sludge production, lb/d
(c) Primary solids contents, percent
(d) Secondary solids content, percent
(e) Specific gravity
(f) Volatile solids content, percent

Table 15.1
Aerobic digester design criteria

Parameter	Value
Hydraulic detention time, days at 20°C[a]	
Activated sludge only	12 to 16
Activated sludge from plant operated without primary settling	16 to 18
Primary plus activated or trickling-filter sludge	18 to 22
Solids loading, lb volatile solids/ft^3/day	0.1 to 0.2
Oxygen requirements	
BOD_5 in primary sludge, lb/lb cell tissue	$\simeq$2
Energy requirements for mixing	
Mechanical aerators, hp/1000 ft^3	0.5 to 1.0
Air mixing, cfm/1000 ft^3	20 to 30
Dissolved oxygen level in liquid, mg/L	1 to 2

[a] Detention times should be increased for temperature below 20°C. If sludge cannot be withdrawn during certain periods (e.g. weekends, rainy weather), additional storage capacity should be provided. Ammonia produced during carbonaceous oxidation is oxidized to nitrate.

6.2. Design Parameters

The current design criteria for aerobic digesters are summarized in Tables 15.1 and 15.2.

6.3. Design Procedure

6.3.1. Calculate Total Quantity of Raw Sludge

$$Q = W_s(100)/(\text{specific gravity})(\text{percent solids})(8.34) \tag{5}$$

where

Q = Quantity of raw sludge production, gpd
W_s = Weight of solids in produced sludge, lb/d

6.3.2. Select Hydraulic Detention Time and Calculate Digesters' Volume

$$V = (t)(Q) \tag{6}$$

where

V = volume of digester, gal
t = hydraulic detention time, d
Q = Quantity of raw sludge production, gpd

6.3.3. Check Volatile Solids Loading

$$L_{vs} = W_{vs}(7.48)/V < (0.1\text{–}0.2) \tag{7}$$

where

L_{vs} = volatile solids loading, lb/ft^3/d
W_{vs} = Weight of volatile solids production, lb/d
V = volume of digester, gal

Table 15.2
US EPA aerobic digester design criteria

Parameters	Design criteria	
Solids residence time required to achieve	Days	Liquid temperature
40 percent volatile solids reduction	108	40°F
	31	60°F
	18	80°F
55 percent volatile solids reduction	386	40°F
	109	60°F
	64	80°F
Oxygen requirements	2.0 pounds of oxygen per pound of volatile solids destroyed when liquid temperature 113°F or less	
	1.45 pounds of oxygen per pound of volatile solids destroyed when liquid temperature greater than 113°F	
Oxygen residual	1. 0 mg/L of oxygen at worst design conditions	
Expected maximum solids concentration achievable with decanting	2.5 to 3.5 percent solids when dealing with a degritted sludge or one in which no chemicals have been added	
Mixing horsepower	Function of tank geometry and type of aeration equipment utilized. Should consult equipment manufacturer	
	Historical values have ranged from 0.5 to 4.0 horsepower per 1,000 cubic feet of tank volume	

(*Source:* US EPA) (9).

6.3.4. Calculate Solids Retention Time

(a) Assume percent destruction of volatile solids: 40 percent is common but it increases with temperature and retention time from approximately 33 to 70 percent.
(b) Calculate solids accumulation per day

$$W_{ac} = W_s - W_s(\%\ \text{volatile}/100)(\%\ \text{destroyed}/100)(0.75) \tag{8}$$

where
W_{ac} = solids accumulation, lb/d
W_s = weight of solids in produced sludge, lb/d

(c) Assume MLSS in digester and calculate total digester capacity

$$W_{dc} = (V)(\text{MLSS})(8.34)(10^{-6}) \tag{9}$$

where
W_{dc} = digester capacity, lb
MLSS = mixed liquor SS in digester, mg/L
V = volume of digester, gal

(d) Calculate solids retention time

$$\text{SRT} = \frac{W_{\text{dc}}}{W_{\text{ac}}} \tag{10}$$

where
SRT = solids retention time, d
W_{dc} = digester capacity, lb
W_{ac} = solids accumulation, lb/d

6.3.5. Calculate Sludge Wasting Schedule

Assume solids content in digested sludge is approximately 2.5 percent.

$$V_{\text{w}} = \frac{\text{total sludge in digester (lb)}(100)}{(\text{specific gravity})(\%\ \text{solids})(8.34)} \tag{11}$$

where
V_{w} = Volume of sludge to be wasted each SRT, gal

6.3.6. Calculate Oxygen Requirement for Bacterial Growth

Assume O_2 required per pound of volatile solids destroyed.

$$O_2 = (O_{\text{r}})W_{\text{s}}(\%\ \text{volatile}/100)(\%\ \text{destroyed}/100) \tag{12}$$

where
O_2 = total oxygen required, lb/d
O_{r} = oxygen required/lb of volatile solids destroyed = 2.0 lb
W_{s} = Weight of solids in produced sludge, lb/d

6.3.7. Calculate Air Requirement for Mixing

(a) Assume standard transfer efficiency, percent
(b) Assume constants α , β , and ρ
(c) Select summer temperature T
(d) Calculate operating transfer efficiency

$$\varepsilon_{\text{o}} = \varepsilon_{\text{s}}[(C_{\text{s}})_T\ \beta\rho - C_{\text{L}}]\,\alpha\,(1.024)^{\text{T}-20}/9.20 \tag{13}$$

where
ε_{o} = operating transfer efficiency, percent
ε_{s} = standard transfer efficiency, percent (5–8%)
$(C_{\text{s}})_{\text{T}}$ = oxygen saturation at the summer temperature
$\beta = (C_{\text{s}}\ \text{waste}/C_{\text{s}}\ \text{water}) = 0.9$
ρ = correction for altitude = 1.0
C_{L} = minimum oxygen to be maintained in the digester, mg/L
$\alpha = (K_{\text{La}}\ \text{waste}/K_{\text{La}}\ \text{water}) = 0.9$
K_{La} = oxygen transfer coefficient
T =temperature, °C

(e) Calculate air supply; check against a minimum of 20 cfm/1000 ft^3.

$$Q_{\text{air}} = O_2(7.48)(10^5)/(\varepsilon_{\text{o}}\%)(0.0176\,\text{lb}\,O_2/\text{ft}^3\ \text{air})\ 1440\ (\text{min}/\text{d})\ V \tag{14}$$

where
Q_{air} = air supply, cfm/1000 ft^3
O_2 = total oxygen required, lb/d
ε_o = operating transfer efficiency, percent
V = volume of the basin, gal

6.4. Output Data

(a) Raw sludge specific gravity
(b) Detention time, days
(c) Volatile solids destroyed, percent
(d) Mixed liquor solids, mg/L
(e) Solids in digested sludge, percent
(f) Rate constant, BOD_5 applied
(g) Coefficient of oxygen saturation in waste/oxygen saturation in water
(h) Standard transfer efficiency, percent
(i) Digester volume, gal
(j) Volatile solids loading, lb/ft^3/d
(k) Solids accumulation, lb/d
(l) Volume of wasted sludge, gal
(m) Solids retention time, d
(n) Oxygen requirement, lb/d
(o) Air supply, cfm/1,000 ft^3

7. COST

For detailed discussion on cost and the choice of cost effective wastewater and biosolids treatment systems, the reader is referred to References (12, 69–72).

7.1. Capital Cost

A regression analysis of construction bids indicated that the capital cost could be approximated by a mathematical relationship (9). The formula has been updated to the year 2008 using U.S. Army Corps of Engineers Civil Works Construction Cost Index System for utilities, (72, Appendix A), and is given in Equation (15).

$$C = 3.77 \times 10^5 Q^{1.14} \tag{15}$$

where
C = capital cost of process in 2008 USD
Q = plant design flow, MGD

The associated costs included those for excavation, process piping, equipment, concrete, and steel.

Table 15.3
Aerobic digestion labor requirement

	Labor, man-h/yr		
Plant design flow, MGD	Operation	Maintenance	Total
0.5	100	20	120
1	160	30	190
2	260	50	310
5	500	100	600
10	800	160	960
25	1,500	300	1,800

Source: US EPA (73).

In addition, such costs as those for administrating and engineering are equal to 0.2264 of the capital cost (9):

$$C_{A\&E} = 0.2264\,C \tag{16}$$

where

$C_{A\&E}$ = Cost of administration and engineering in 2008 USD

7.2. Operation and Maintenance Cost

Although there are many items that contribute to operation and maintenance cost, in most aerobic digestion systems, the two most prevalent are staffing requirements and power usage.

7.2.1. Staffing Requirements

Table 15.3 lists labor requirements for both operation and maintenance (73). The labor indicated includes: checking mechanical equipment, taking dissolved oxygen and solids analyses, and general maintenance around the clarifier.

7.2.2. Power Requirements

The cost of power for operating aeration equipment has become a significant factor. It is possible to minimize power consumption through two developments in environmental science (74).

(a) Make sure that the tank geometry and aeration equipment are compatible. The difference between optimized and unoptimized design can mean as much as a 50 percent difference in power consumption.
(b) Use devices to control oxygen (power) input. Because of temperature effects, oxygen requirements for any given aerobic digestion system can vary as much as 20 to 30 percent between summer and winter. One must design to meet the worst conditions (summer), for without some type of oxygen controller, considerable power is wasted during other times of the year.

7.2.3. Other Requirements

In addition to manpower and power cost, the designer must consider lubrication requirements. If mechanical aerators are being used, each unit needs to have an oil change once, and preferably twice, a year. Depending on horsepower size, this could be 5 to 40 gallons per unit per change (19–152 L/unit/change). Further, the designer must make sure an adequate inventory of spare parts is available.

8. RECENT DEVELOPMENTS AND SUMMARY

8.1. Recent Developments

There have been many new developments in both aerobic and anaerobic digestion processes since 1990 (75–90). Detailed new developments on anaerobic digestion may be found in Ref. 90. The advantages of using combined aerobic digestion and anaerobic digestion are reported by many researchers (75, 76, 79–81, 87–90). This chapter will introduce two new developments in aerobic digestion:

(a) Vertical shaft digestion; and
(b) Cryophilic aerobic digestion.

8.1.1. Vertical Shaft Digestion

There are two types of autothermal thermophilic aerobic digestion (ATAD) processes: (a) ATAD using air known as ATAD-Air process, and (b) ATAD using pure oxygen known as ATAD-Oxygen process. The new vertical shaft digestion (VSD) process can be either VSD-ATAD-Air process involving the use of a vertical shaft reactor, or VSD-ATAD-Oxygen process involving the use of a vertical shaft reactor. A vertical shaft reactor is typically 350 ft in depth and 2.5 to 10 ft in diameter.

Both VSD-ATAD-Air and VSD-ATAD-Oxygen configurations with vertical shaft reactors are designed and marketed as VERTAD systems by NORAM Engineering and Constructors, Ltd, Vancouver, BC, Canada (see Figure 15.7). The principal difference between VERTAD and conventional ATAD-Air and ATAD-Oxygen systems is its inground hyperbolic vertical shaft reactor for aerobic digestion. Installed by conventional drilling techniques, the VERTAD's 350-ft deep vertical shaft reactor occupies only a fraction of the area used by conventional surface digestion systems (87). The injected air, in addition to supplying the required oxygen, also provides airlift and circulation, eliminating the need for any mechanical mixing. The vertical shaft reactor can be operated in batch mode or with continuous feed. The reactor is designed with a plug-flow zone at the bottom of the shaft before the treated biosolids are discharged from the system. This zone prevents any short-circuiting, providing the high-temperature residence time required for meeting Class A biosolids requirements (87).

The depth of the vertical shaft reactor accompanied by the high pressure enables the system to attain its high oxygen transfer efficiency (over 40%). This high rate of oxygen transfer enables higher metabolic activity and greater than 40% volatile solids destruction in a retention time shorter than 4 days (87). In addition, the pressurized nature of the VERTAD system (either VSD-ATAD-Air or VSD-ATAD-Oxygen) enables easy flotation thickening of the product to 8–12% solids with dissolved gasses from the process. The new flotation process

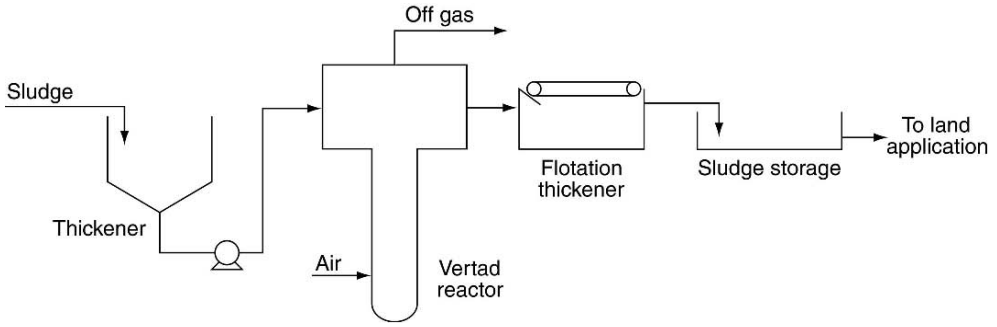

Fig. 15.7. Vertical shaft digestion (VSD) process system (VERTAD) (*Source:* NORAM Engineering and Constructors Ltd., Vancouver, Canada).

involving the use of vertical shaft reactor is termed vertical shaft flotation. The thickened product dewaters to over 30% solids with a relatively low polymer demand (approximately 14 lb/ton).

Advantages of the higher operating temperatures in the VSD-ATAD-Air and the VSD-ATAD-Oxygen systems include:

(a) Reduced HRT (3–6 days) for obtaining 35–45% volatile solids reduction
(b) Heat is generated that can be recovered for building and/or process heating (preheating sludge, heating subsequent sludge treatment processes), and
(c) Pasteurization, resulting in Class A biosolids production

Reported disadvantages of VSD-ATAD-Air and VSD-ATAD-Oxygen systems include:

(a) High power costs associated with aeration (which have decreased in newer generation VSD-ATAD systems that achieve higher oxygen transfer)
(b) Higher polymer costs associated with product dewatering.

8.1.2. Cryophilic Aerobic Digestion

Cryophilic digestion involves the operation of aerobic digestion systems in lower temperature ranges (below 20°C). This mode of digestion is particularly relevant in some treatment facilities in countries with colder climate. Longer solids retention times are required at lower temperatures. It has been reported that at low temperatures (5–20°C) a processing time of 250 to 300 degree-days is required to achieve reasonable volatile solids destruction (67, 89).

8.2. Summary

There are seven primary variations of the aerobic digestion process:

1. Conventional aerobic digestion using air
2. Conventional aerobic digestion using oxygen
3. Auto thermal thermophilic aerobic digestion using air (ATAD-Air)
4. Auto thermal thermophilic aerobic digestion using oxygen (ATAD-Oxygen)
5. ATAD-Air process using vertical shaft reactor for aeration/digestion (VSD-ATAD-Air)

6. ATAD-Oxygen process using vertical shaft reactor for oxidation/digestion (VSD-ATAD-Oxygen), and
7. Cryophilic aerobic digestion

The theory and principles of all seven aerobic digestion processes are alike. Each of the processes is a suspended-growth biological treatment process for the stabilization of the biosolids produced in wastewater treatment facilities (75–99). Wastewater biosolids (both primary and secondary waste-activated sludge) are stabilized by the destruction of the biodegradable organic components and the reduction of pathogenic organisms. Aerobic digestion is based upon endogenous respiration, where in the absence of suitable substrate food, microorganisms begin to digest their own protoplasm to obtain energy (87, 90, 92–99). Cell tissue is aerobically oxidized to carbon dioxide, water, and ammonia or nitrates. Some of the energy released by the microbial degradation is used to form new cellular material, but the majority is released as heat; thus the aerobic oxidation process is exothermic.

Advantages claimed for all aerobic digestion processes are as follows:

1. Relatively simple operation with volatile solids reduction slightly less than anaerobic systems
2. Low BOD concentrations in the effluent supernatant
3. Production of an odorless, biologically stable end product
4. Recovery of more of the sludge fertilizer value, and
5. Reduced capital costs

Reported disadvantages of all aerobic digestion processes include:

1. High power costs associated with aeration
2. Higher polymer costs associated with product dewatering, and
3. High cost associated with oxygen generation in the case of high-purity oxygen digestion (83, 89)

Vertical shaft digestion (also known as VERTAD by NORAM Engineering and Constructors Ltd. of Vancouver, Canada) is a newer generation of ATAD-Air and ATAD-Oxygen processes involving the use of vertical shaft reactors for oxidation/digestion. The vertical shaft reactor is typically 350 ft in depth, and 2.5 to 10 ft. in diameter, offer extremely high oxygen transfer efficiency and extremely small foot print for construction. The HRT can be shortened to 3 to 6 days and power costs can be reduced in comparison with conventional ATAD-Air or ATAD-Oxygen process.

Cryophilic aerobic digestion is not a very new process, because aerobic digestion has been used in northern cold climates since 1970. It has been given the new name (67, 89), and its design criteria has been preliminarily established.

9. DESIGN EXAMPLES

9.1. Example 1

A wastewater treatment plant produces sludge with the following characteristics:

1. Solids content = 2,191 lb/d.
2. Specific gravity = 1.05

3. Solids concentration = 1.0%
4. Volatile solids = 80%

Design an aerobic digester for stabilizing the sludge by going through the following steps:

First: Calculate total quantity of raw sludge
Second: Select hydraulic detention time and calculate volume of digester
Third: Check volatile solids loading
Fourth: Calculate solids retention time
Fifth: Calculate sludge wasting schedule
Sixth: Calculate oxygen requirement for bacterial growth
Seventh: Calculate air requirement for mixing

Solution

First: Calculate total quantity of raw sludge

$$Q = \frac{W_s(100)}{(\text{specific gravity})\ (\text{percent solids})\ (8.34)} \tag{5}$$

where

Q = Quantity of raw sludge production, gpd
W_s = Weight of solids in produced sludge, lb/d = 2,191 lb/d

$$Q = 2,\,191(100)/(1.05)(1.0)(8.34)$$
$$Q = 25,\,020\,gpd$$

Second: Select hydraulic detention time and calculate volume of digester

$$V = (t)(Q) \tag{6}$$

where

V = volume of digester, gal
t = hydraulic detention time = 15 d
Q = Quantity of raw sludge production = 25,020 gpd

$$V = (15)(25,\,020)$$
$$V = 375,\,300\,gal$$

Third: Check volatile solids loading

$$\mathrm{L_{vs}} = \frac{W_{vs}(7.48)}{V} < (0.1\text{–}0.2) \tag{7}$$

where

L_{vs} = volatile solids loading, lb/ft^3/d
W_{vs} = Weight of volatile solids production = 2,191 (0.80) = 1,753 lb/d
V = volume of digester = 375,300 gal

$$L_{vs} = 1{,}753(7.48)/375{,}300$$
$$L_{vs} = 0.035 < 0.1\ \text{OK}$$

Fourth: Calculate solids retention time

(a) Assume percent destruction of volatile solids = 50%
(b) Calculate solids accumulation per day

$$W_{ac} = W_s - W_s(\%\ \text{volatile}/100)(\%\ \text{destroyed}/100)(0.75) \tag{8}$$

where
W_{ac} = solids accumulation, lb/d
W_s = weight of solids in produced sludge = 2,191 lb/d

$$W_{ac} = 2{,}191 - 2{,}191(80/100)(50/100)0.75$$
$$W_{ac} = 1{,}534\ \text{lb/d}$$

(c) Assume MLSS in digester and calculate total digester capacity

$$W_{dc} = (V)(\text{MLSS})(8.34)(10^{-6}) \tag{9}$$

where
W_{dc} = digester capacity, lb
MISS = mixed liquor SS in digester = 12,000 mg/L
V = volume of digester = 375,300 gal

$$W_{dc} = 375{,}300(12{,}000)(8.34)(10^{-6})$$
$$W_{dc} = 37{,}560\ \text{lb}$$

(d) Calculate solids retention time

$$\text{SRT} = \frac{W_{dc}}{W_{ac}} \tag{10}$$

where
SRT = solids retention time, d
W_{dc} = digester capacity = 37,560 lb
W_{ac} = solids accumulation 1,534 lb/d

$$\text{SRT} = 37,560/1,534$$
$$\text{SRT} = 24.5\,d$$

Fifth: Calculate sludge wasting schedule
Assume solids content in digested sludge is approximately 2.5 percent.

$$V_w = \frac{\text{total sludge in digester (lb)}(100)}{(\text{specific gravity})(\%\ \text{solids})(8.34)} \tag{11}$$

where
V_w = Volume of sludge to be wasted each SRT, gal

$$V_w = 37{,}560(100)/1.05(2.5\%)(8.34)$$
$$V_w = 171,566\ \text{gal to be wasted every 24.5 days (about 7,000 gal/d)}$$

Sixth: Calculate oxygen requirement for bacterial growth
Assume O_2 required per pound of volatile solids destroyed = 2.0 lb.

$$O_2 = (O_r)W_s(\%\ \text{volatile}/100)(\%\ \text{destroyed}/100) \tag{12}$$

where

O_2 = total oxygen required, lb/d
O_r = oxygen required/lb of volatile solids destroyed = 2.0 lb
W_s = Weight of solids in produced sludge = 2,191 lb/d

$$O_2 = 2(2,191)(0.80)(0.50)$$
$$O_2 = 1,753\ \text{lb/d}$$

Seventh: Calculate air requirement for mixing.

(a) Assume standard transfer efficiency = 8%
(b) Assume constants α, β, and ρ

$$\alpha = 0.9$$
$$\beta = 0.9$$
$$\rho = 1.0$$

(c) Select summer temperature T = 25°C
(d) Calculate operating transfer efficiency

$$\varepsilon_o = \varepsilon_s[(C_s)_T\ \beta\rho - C_L]\alpha(1.024)^{T-20}/9.20 \tag{13}$$

where

ε_o = operating transfer efficiency, %
ε_s = standard transfer efficiency = 8%
α = (K_{La} waste/K_{La} water) = 0.9
β = (C_s waste/C_s water) = 0.9
ρ = correction for altitude = 1.0
C_L = minimum oxygen to be maintained in the digester = 2.0 mg/L
K_{La} = oxygen transfer coefficient
T = temperature = 25°C
$(C_s)_T$ = oxygen saturation at the summer temperature = 8.2 mg/L

$$\varepsilon_o = 8[(8.2)(0.9)(1.0) - 2.0](0.9)(1.024)^{25-20}/9.20$$
$$\varepsilon_o = 4.7\%$$

(e) Calculate air supply; check against a minimum of 20 cfm/l000 ft^3.

$$Q_{air} = O_2(7.48)(10^5)/(\varepsilon_o\%)(0.0176\ \text{lb}\ O_2/\text{ft}^3\ \text{air})1440\ (\text{min/d})\ V \tag{14}$$

where

Q_{air} = air supply, cfm/1000 ft^3
O_2 = total oxygen required = 1,753 lb/d
ε_o = operating transfer efficiency = 4.7%
V = volume of the basin = 375,300 gal

$$Q_{air} = 1,753(7.48)(10^5)/(4.7\%)(0.0176\ \text{lb}\ O_2/\text{ft}^3\ \text{air})1440\ (\text{min/d})375,300$$
$$Q_{air} = 29.3\ \text{cfm}/1,000\ \text{ft}^3 > 20\ \text{OK}$$

9.2. Example 2

A design engineer has determined that the following quantities of sludge will be produced at a 0.5 MGD (22 L/s) contact stabilization plant:

(a) Total daily solids generation = 1,262 lb (572 kg)
(b) Amount due to chemical sludge = 0
(c) Amount that will be volatile = 985 lb (447 kg)
(d) Amount that will be non-volatile = 277 lb (125 kg)

In addition, the designer has the following information:

(a) Estimated minimum liquid temperature (winter) in digester is 50°F (10°C).
(b) Estimated maximum liquid temperature (summer) in digester is 77°F (25°C).
(c) System must achieve greater than 40 percent volatile solids reduction during the winter.
(d) A minimum of two continuously operated tanks are required. This is a state requirement for plants under 1 MGD (44 L/s).
(e) Expected waste sludge solids concentration to the aerobic digester is 8,000 mg/L.
(f) Expected thickened solids concentration for the stabilized sludge is three percent (30,000 mg/L), based on designer's experience.

It is required to design the digestion system by determining the following:

(a) Sludge age required
(b) Amount of volatile solids reduction
(c) Oxygen requirements
(d) Tank Volume
(e) Power requirement and tanks dimensions
(f) Clarifier surface area
(g) Supernatant flow

Solution

(a) Sludge Age Required
Figure 15.6 offers a quick method for calculating the number of degree days required to achieve the 40 percent volatile solids reduction required. The result is 475 degree-days.

At a basin temperature of 50°F (10°C) then:

$$475 \text{ degree-days}/10 \text{ degrees} = 47.5 \text{ d}$$

Therefore, the volume of the aerobic digester must be adequate to provide 47.5 days sludge age to meet minimum volatile solids reduction during the winter.

During the summer, the basin temperature will be 77°F (25°C):

$$25°\text{C} \times 47.5 \text{ day sludge age} = 1{,}175 \text{ degree-days.}$$

From Figure 15.6, at 1,175 degree-days, there would be 49 percent volatile solids reduction.

(b) Amount of volatile Solids Reduction
For winter conditions, there would be a 40 percent volatile solids (VS) reduction. The actual pounds of solids reduced are:

$$985 \text{ lb VS/d} \times 0.4 = 394 \text{ lb VS reduced/d } (179 \text{ kg/d})$$

For summer conditions, there would be a 49 percent volatile solids reduction. The actual pounds of solids reduced are:

$$985\,\text{lb VS/d} \times 0.49 = 483\,\text{lb VS reduced/d}\ (219\,\text{kg/d})$$

(c) Oxygen Requirements
Since nitrification is expected, provisions must be made to supply 2.0 lb of oxygen/lb of volatile solids destroyed (2 kg O_2/kg volatile solids destroyed).

Winter conditions:

$$394\,\text{lb VS destroyed/d} \times 2.0\,\text{lb } O_2\text{/lb VS destroyed} = 788\,\text{lb } O_2\text{/d}\ (358\,\text{kg/d})$$

Summer conditions:

$$483\,\text{lb VS destroyed} \times 2.0\,\text{lbs } O_2\text{/lb VS destroyed} = 966\,\text{lb } O_2\text{/d}\ (438\,\text{kg/d})$$

During summer conditions, a minimum of 1.0 mg/L oxygen residual must be provided.

(d) Tank Volume
Sludge age in an aerobic digester can be defined as follows:

Sludge age = total lb SS in aerobic digester / total lb SS lost per day from aerobic digester

The suspended solids concentration in the digester will range from the value of the influent suspended solids concentration or 8,000 mg/L to the maximum value of the thickened and stabilized solids concentration of 30,000 mg/L. On the average, the suspended solids concentration within the digester is equal to 70 percent of the thickened solids concentration, or $30{,}000 \times 0.70 = 21{,}000$ mg/L.

An average poundage of suspended solids in the supernatant can be approximated by the following equation:

(SS concentration in supernatant) (l-f)(8.34)(influent flow)

where
f = the fraction of influent flow into the aerobic digester that is retained, and
1-f = the fraction that leaves as supernatant.
The term f can be approximated by the following equation:

f = (influent SS concentration/thickened SS concentration) (fraction of solids not destroyed)

For winter conditions, the fraction of solids not destroyed is:

$$(1{,}262\,\text{lb total solids} - 394\,\text{lb of solids reduced})/1{,}262\,\text{lb total solids} = 0.69$$

Then, the term f for this example is:

$$(8{,}000\,\text{mg/L}/30{,}000\,\text{mg/L}) \times 0.69 = 0.18$$

Therefore, 18 percent of the influent flow into the aerobic digester will be retained, and 82 percent will leave as supernatant.

For a properly designed solids-liquid separator (under 200 gpd/ft^2 = 8.16 m^3/d/m^2 overflow rate), the suspended solids concentration would be approximately 300 mg/L.

The influent flow can be found by dividing the influent solids load (1,262 lb/d = 572 kg/d) by the influent solids concentration (8000 mg/L). The result is 18,914 gpd (71.5 m^3/d).

The pounds of suspended solids intentionally wasted per day from the aerobic digestion system can now be approximated from the following expression:

(SS concentration in thickened sludge) (f)(8.34)(influent flow)

It is now possible to solve for the required tank volume, V, for any given sludge age. In this example, winter conditions govern, and it was previously calculated that a time of 47.5 days minimum was required. From the values previously discussed:

$$47.5\,\text{d} = (21{,}000\,\text{mg/L})(8.34)(\text{V MG})/[300\,\text{mg/L}(1 - 0.18) + (30{,}000)(0.18)](8.34)(0.0189\,\text{MG})$$

$$\text{Tank volume, V} = 0.233\,\text{MG}(881\,\text{m}^3)$$

Theoretical hydraulic detention time:

$$233{,}000\,\text{gal}/18{,}915\,\text{gpd} = 12.3\,\text{d}$$

This is the minimum volume, to which must be added capacity for weekend storage and precipitation requirements. For this design, two tanks will be provided, each to have a volume capacity of 233,000 gal (881 m^3) (100 percent stand-by capacity as per state requirements).

(e) Power requirement and tanks dimensions

Select a depth of 12 ft (3.65m). Since each tank is to be 233,000 gal (881 m^3), the surface area with a 12-ft (3.65 m) water depth would be:

$$\text{Area} = \text{V/depth}$$

$$\text{Area} = 233,000/7.48(12)$$

$$= 2596\,\text{ft}^2\,(241\,\text{m}^2)$$

The aerator manufacturer recommends that a minimum 10 horsepower unit (7.5 kw) will be required to mix the 12-ft (3.65 m) liquid depth. Each 10 horsepower unit (7.5 kw) could mix an area 40 ft by 40 ft (12.1 m by 12.1 m). After making some calculations, it is found that there is a need for two 10-horsepower (7.5 kw) units in each tank, each tank being 36 ft (10.9 m) wide by 72 ft (24.5 m) long and having a total tank depth of 14 ft (4.2 m) allowing 2 ft (0.61 m) of free board. Figure 15.8 shows a view of the plan.

(f) Clarifier Surface Area

Surface area was based on an overflow rate of 200 gal/ft^2/d (8.16 $\text{m}^3/\text{d}/\text{m}^2$). At an influent flow of 18,915 gal/d (71.5 m^3/d), the required surface area is:

$$\text{Area} = 18{,}915/(200) = 95\,\text{ft}^2\,(8.8\,\text{m}^2).$$

Using a circular clarifier:

$$\text{Diameter}^2 = [95(4/3.14)] = 121$$

$$\text{Diameter} = 11\,\text{ft}$$

Select a 12-foot (3.7 m) diameter clarifier.

(g) Supernatant Flow

It was previously calculated that 82 percent of the influent to the aerobic digester would leave as supernatant. Based on an influent of 18,915 gal/d (71.5 m^3/d), the supernatant flow will be:

$$18{,}910(0.82) = 15{,}510\,\text{gal/d}\,(58.6\,\text{m}^3/\text{d}),\ \text{plus any precipitation.}$$

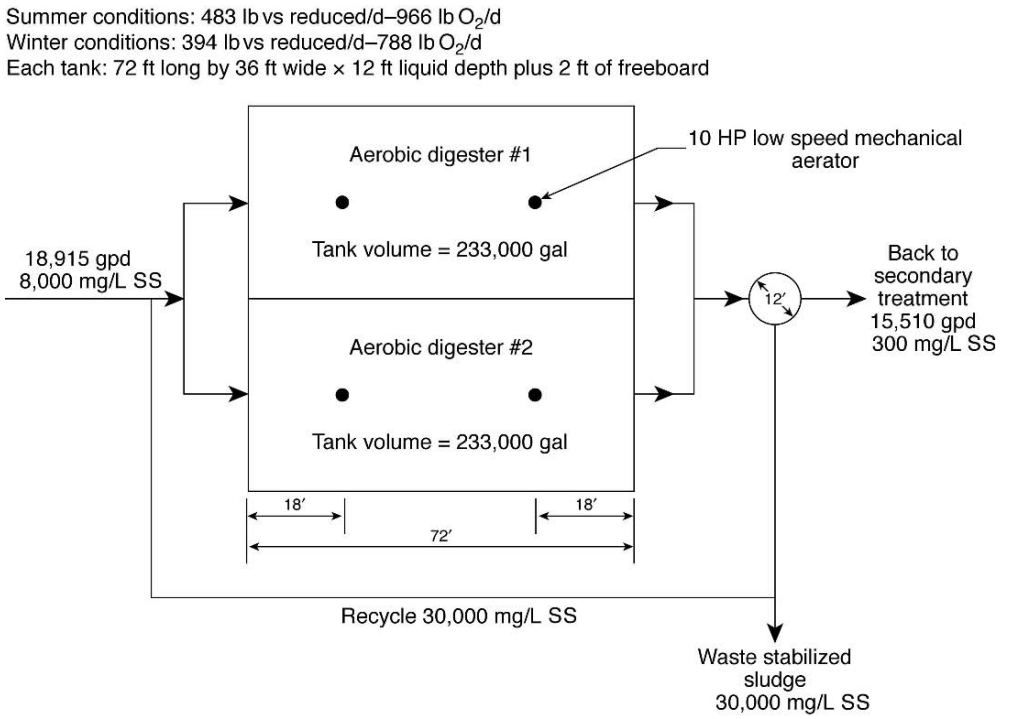

Fig. 15.8. Aerobic digestion design Example 2 (*Source:* US EPA) (9).

NOMENCLATURE

$\alpha = (K_{La}\,\text{waste}/K_{La}\,\text{water}) = 0.9$
$\beta = (C_s\,\text{waste}/C_s\,\text{water}) = 0.9$
C = capital cost of process in 2008 USD
$C_{A\&E}$ = Cost of administration and engineering in 2008 USD
$(C_s)_T$ = oxygen saturation at the summer temperature, T
C_L = minimum oxygen to be maintained in the digester, mg/L
dM/dt = rate of change of biodegradable volatile solids per unit of time, mg/L/d
ε_o = operating transfer efficiency, percent
ε_s = standard transfer efficiency, percent (5–8%)
K_d = reaction rate constant, d^{-1}
K_{La} = oxygen transfer coefficient
L_{vs} = volatile solids loading, $lb/ft^3/d$
M = concentration of biodegradable volatile solids remaining at time t in the aerobic digester, mg/L (ppm)
MLSS = mixed liquor SS in digester, mg/L
O_2 = total oxygen required, lb/d

O_r = oxygen required/lb of volatile solids destroyed = 2.0 lb
ρ = correction for altitude = 1.0
Q = plant design flow, MGD
Q = Quantity of raw sludge production, gpd
Q_{air} = air supply, cfm/1000 ft^3
SRT = solids retention time, d
t = time, d
t = hydraulic detention time, d
T = temperature, °C
V = volume of digester, gal
V = volume of basin, gal
V_w = Volume of sludge to be wasted each SRT, gal
W_{ac} = solids accumulation, lb/d
W_{dc} = digester capacity, lb
W_s = Weight of solids in produced sludge, lb/d
W_{vs} = Weight of volatile solids production, lb/d

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APPENDIX

United States Yearly Average Cost Index for Utilities*

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

*Extracted from U.S. ACE (2000-Tables Revised 31 March 2003) *Civil Works Construction Cost Index System Manual*, # 1110-2-1304, U.S. Army Corps of Engineers, Washington, DC, USA, PP 44 (PDF file is available on the Internet at http://www.nww.usace.army.mil/cost (72).

16
Biosolids Composting

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract Composting is one of several methods for treating biosolids to create a marketable end product that is easy to handle, store, and use. The end product is usually a Class A, humus-like material without detectable levels of pathogens that can be applied as a soil conditioner and fertilizer to gardens, food and feed crops, and rangelands. This compost provides large quantities of organic matter and nutrients (such as nitrogen and potassium) to the soil, improves soil texture and elevates soil cation exchange capacity. Biosolids composting is the aerobic thermophilic decomposition of organic constituents to a relatively stable humus-like material. Environmental factors influence the activities of bacteria, fungi and actinomycetes in this decomposition process and affect the speed and course of composting cycles. The volatility and type of material, moisture content, oxygen concentration, carbon/nitrogen ratio, temperature, and pH are key determinants in the process. Process applicability, compost quality, design criteria, windrow process, aerated static pile process, in-vessel composting, costs and design examples are discussed.

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Key Words Composting • Class A biosolids • fertilizers • windrow • aerated static pile • in-vessel composting.

1. INTRODUCTION

Composting is one of several methods for treating biosolids to create a marketable end product that is easy to handle, store, and use. The end product is usually a Class A, humus-like material without detectable levels of pathogens that can be applied as a soil conditioner and fertilizer to gardens, food and feed crops, and rangelands. This compost provides large quantities of organic matter and nutrients (such as nitrogen and potassium) to the soil, improves soil texture, and elevates soil cation exchange capacity (an indication of the soil's ability to hold nutrients), all characteristics of a good organic fertilizer. Biosolids compost is safe to use and generally has a high degree of acceptability by the public. Thus, it competes well with other bulk and bagged products available to homeowners, landscapers, farmers, and ranchers (1).

Since the early 1900s biosolids have been composted as a minor constituent of refuse in many countries. However, only since the early seventies has increased attention been directed to composting of wastewater biosolids as an environmentally sound alternative to stabilization for biosolids reuse or disposal.

A major study of the composting of wastewater biosolids was conducted at Salt Lake City from 1967 to 1969 (2). This work was followed in 1972 by research at pilot-scale wastewater biosolids composting facilities at the USDA Agricultural Research Center at Beltsville, Maryland (3–5) and full-scale operations at County Sanitation Districts of Los Angeles County plant at Carson, California. Based on the operating experiences and developments at these plants, new projects were undertaken at Bangor, Maine (6); Durham, New Hampshire (7); and Windsor, Ontario (8).

Since 1984, US EPA has encouraged the beneficial use of wastewater residuals through formal policy statements. The implementation of Part 503 Rule enhanced the acceptance of biosolids as a resource by standardizing metal and pathogen concentrations. Moreover, Part 503 Rule officially identifies composting as a method to control pathogens and reduce vector attraction (9).

In 1983 the number of composting facilities in the U.S. numbered only 61 but by 1988 the number of biosolids composting plants has risen to 115 (10, 11). According to a 1998 survey in *Biocycle, The Journal of Composting and Recyclin*g, 274 biosolids composting facilities were operating in the United States (12). Nearly 50 additional facilities were in various stages of planning, design, and construction. A large number of these facilities (over 40 percent) use the aerated static pile composting method.

Biosolids composting is the aerobic thermophilic decomposition of organic constituents to a relatively stable humus-like material (13). Environmental factors influence the activities of bacteria, fungi, and actinomycetes in this decomposition process and affect the speed and course of composting cycles. The volatility and type of material, moisture content, oxygen concentration, carbon/nitrogen ratio, temperature, and pH are key determinants in the process (14). Biosolids are not rendered totally inert by composting. The composting process is

considered complete when the product can be stored without giving rise to nuisances such as odors, and when pathogenic organisms have been reduced to a level such that the material can be handled with minimum risk.

Processes for composting wastewater biosolids differ from those for composting refuse. There are several principal advantages of biosolids composting as compared to refuse composting (15):

1. Biosolids composting does not require the complex materials management and separation techniques necessary for most refuse composting operations.
2. Municipal wastewater biosolids are more uniform in composition causing less operating difficulties.
3. The final composted mixture utilizing biosolids is more suitable for marketing because it generally does not contain the plastics, metal, and glass commonly found in refuse compost.
4. Biosolids composting is often viewed as an alternative disposal method and does not have to be evaluated on profit-making potential as some refuse composting operations have been.

There are three general methods of composting biosolids:

1. Windrow
2. Aerated static pile, and
3. In-vessel systems

Each method uses the same scientific principals but varies in procedures and equipment needs. The first two processes are not enclosed, although a roof may be provided to protect the compost from precipitation. Both processes make use of portable mechanical equipment such as front-end loaders or mixers for compost mixing and turning. In-Vessel systems utilize a stationary-enclosed vessel or reactor for mechanical composting.

2. APPLICABILITY AND ENVIRONMENTAL IMPACT

Biosolids composting has grown in popularity for the following reasons (16):

1. Lack of availability of landfill space for solids disposal.
2. Composting economics are more favorable when landfill tipping fees escalate.
3. Emphasis on beneficial reuse at federal, state, and local levels.
4. Ease of storage, handling, and use of composted product.
5. Addition of biosolids compost to soil increases the soil's phosphorus, potassium, nitrogen, and organic carbon content.

Compost produced from municipal wastewater biosolids can provide a portion of the nutrient requirements for growth of crops. The organic matter in compost is particularly beneficial as a soil conditioner, because it has been stabilized, decomposes slowly, and remains effective for a longer time than the organic matter in uncomposted wastes. Composted biosolids can improve the quality of soils containing excessive amounts of sand or clay as well as already more balanced soils. The use of biosolids compost as a soil conditioner results in the following environmental benefits (17–19):

1. The recycling of a valuable resource
2. Reduction of dependence on chemical fertilizers

3. Offsetting the use of natural resources such as trees or peat moss as mulch material
4. Provides organic nitrogen, phosphorus, and potassium
5. Provides essential plant micronutrients
6. Can reduce the need for pesticides
7. Increased water holding of soils
8. Increased aeration and drainage for clay soils
9. Increased permeability for clay soils
10. Greater root depth
11. Increased microbial population
12. Decreased surface crusting of soils

Composted biosolids can also be used in various land applications. Compost mixed with appropriate additives creates a material useful in wetland and mine land restoration. The high organic matter content and low nitrogen content common in compost provides a strong organic substrate that mimics wetland soils, prevents overloading of nitrogen, and adsorbs ammonium to prevent transport to adjacent surface waters. Compost amended strip-mine spoils produce a sustainable cover of appropriate grasses, in contrast to inorganic-only amendments which seldom provide such a good or sustainable cover (20).

Compost-enriched soil can also help suppress diseases and ward off pests. These beneficial uses of compost can help growers save money, reduce use of pesticides, and conserve natural resources. Compost also plays a role in bioremediation of hazardous sites and pollution prevention. Compost has proven effective in degrading or altering many types of contaminants, such as wood-preservatives, solvents, heavy metals, pesticides, petroleum products, and explosives. Some municipalities are using compost to filter stormwater runoff before it is discharged to remove hazardous chemicals picked up when stormwater flows over surfaces such as roads, parking lots, and lawns. Additional uses for compost include soil mulch for erosion control, silviculture crop establishment, and sod production media (21).

On the negative side, biosolids composting may include the following disadvantages (17, 18):

1. Odor production at the composting site.
2. Survival and presence of primary pathogens in the product.
3. Dispersion of secondary pathogens such as *Aspergillus fumigatu*s, particulate matter, other airborne allergens.
4. Lack of consistency in product quality with reference to metals, stability, and maturity.

Odors from a composting operation can be a nuisance and a potential irritant. Offensive odors from composting sites are the primary source of public opposition to composting and have led to the closing of several otherwise well-operated composting facilities. Although research shows that biosolids odors may not pose a health threat, odors from processing facilities have decreased public support for biosolids recycling programs. Many experts in the field of biosolids recycling believe that biosolids generating and processing facilities have an ethical responsibility to control odors and protect nearby residents from exposure to malodor.

Composting odors are caused by ammonia, amine, sulfur-based compounds, fatty acids, aromatics, and hydrocarbons (such as terpenes) from the wood products used as bulking agents

(22). A properly designed composting plant, such as the one shown in Figure 16.4, operated at a high positive redox potential (highly aerobic) will reduce, but not necessarily eliminate, odors and odor causing compounds during the first 10 to 14 days of the process (23, 24).

In addition to odors, other bioaerosols, such as pathogens, endotoxins, and various volatile organic compounds, must also be controlled. Biofilters are often used to control odors, but the biofilters themselves can give off bioaerosols.

Pathogens, such as bacteria, viruses, and parasites (helminth and protozoa) are present in untreated wastewater residuals. These organisms can potentially invade a normal, healthy human being and produce illness or debilitation. Composting reduces bacterial and viral pathogens to non-detectable levels if the temperature of the compost is maintained at greater than 55°C for 15 days or more (25–27). Additionally, it has been demonstrated that viruses and helminth ova do not regrow after thermal inactivation (25).

Regrowth of *Salmonella* sp. in composted biosolids is a concern, although research shows that salmonellae reach a quick peak during regrowth, then die off. Composting is not a sterilization process and a properly composted product maintains an active population of beneficial microorganisms that compete against the pathogenic members. Under some conditions, explosive regrowth of pathogenic microorganisms is possible. A stabilized product with strict control.of post-composting handling and addition of amendments coupled with four to six weeks of storage will mitigate *Salmonella* regrowth (25).

Compost workers may be exposed to a common fungus known as *Aspergillus fumigatu*s, endotoxins, or other allergens. *A. fumigatus* is common in decaying organic matter and soil. Inhalation of its airborne spores causes skin rashes and burning eyes. While healthy individuals may not be affected, immunocompromised individuals may be at risk. The spores of *A. fumigatus* are ubiquitous and the low risk of exposure is not a significant health concern. However, spore counts at composting facilities are high, and the risk of operators and persons handling composted biosolids being exposed to these spores is also high (23). Inhalation of spores, particulates, and other matter can be reduced or prevented by (23):

1. Wearing masks and other protective devices.
2. Equipping front end loaders with filters or air conditioners.
3. Thoroughly ventilating composting halls.
4. Installing biofilters or other odor scrubbing systems in composting halls.

Organic dust (such as pollen) is another nuisance that must be controlled at composting operations. These contaminants are primarily a concern to workers at the composting facilities and are generally not present in quantities that would cause reactions in most individuals that are not exposed outside of the facilities.

A final point to mention here is that excess nitrogen is detrimental to soil, plants, and water, so care must be taken when choosing application sites, selecting plant/crop types, and calculating the agronomic rate for biosolids land application. It should be noted that the most plant-available form of nitrogen in biosolids (ammonium ion (NH_4^+) is converted to nitrate (NO_3^-) by the composting process. Improper use of biosolids can result in the contamination of water resources with leached nitrogen, because nitrate is more mobile than ammonium, and is taken up less easily by plants. However, applying compost in accordance with the

Part 503 Regulations (9) poses little risk to the environment or public health (28). In fact, the use of compost can have a positive impact on the environment in addition to the soil improving characteristics previously discussed. Reduced dependence on inorganic fertilizers can significantly decrease nitrate contamination of ground and surface waters often associated with use of inorganic fertilizers.

3. COMPOST QUALITY

The persistence of organic chemicals, pathogenic organisms, or heavy metals in some composted biosolids may restrict the use of the material for application to crops for human consumption (13, 29). The U.S. Environmental Protection Agency's 40 CFR Part 503, *Standards for the Use and Disposal of Sewage Sludg*e (9), defines two types of biosolids with respect to pathogen reduction: Class A and Class B. The difference is defined by the degree of pathogen reduction on the solids. When federal performance standards are met, composting insures full destruction of pathogens to *non-detectable levels* in the wastewater solids (i.e., to Class A standards).

The length of time biosolids are composted at a specific temperature is important in determining the eventual use of the compost end product. 40 CFR Part 503 (9), defines time and temperature requirements for both Class A and Class B products (Table 16.1). The production of a Class B product is not always economically justified since the product cannot be used without restrictions and the additional expense to reach Class A requirements can be marginal (1).

In addition to performance standards for the composting process, the Part 503 Rule established maximum concentrations for nine metals which cannot be exceeded in biosolids products, including compost. These are known as ceiling concentrations. The federal maximum allowable metals concentrations are provided in Table 16.2. The Part 503 Rule also established more stringent pollutant concentrations. Biosolids products which do not exceed pollutant concentrations, meet Class A pathogen reduction requirements, and are processed to reduce vector attraction potential are often referred to as *Exception Quality (EQ)* products. Class A and EQ biosolids typically have greater marketing success than Class B biosolids. Control of industrial waste streams to wastewater treatment plants (through pretreatment programs) greatly reduces the presence of metals in pre-processed wastewater residuals, enabling compost to meet the stringent EQ standards of Part 503.

Table 16.1
Time & temperature requirements for biosolids composting (1, 9)

Product	Regulatory Requirements
Class A	**Aerated static Pile or in-vessel**: 55°C for at least 3 days
	Windrow: 55°C for at least 15 days with 5 turns
Class B	40°C or higher for five days during which temperature exceed 55°C for at least four hours

Table 16.2
Maximum metal concentrations in biosolids (9, 18)

Metal	Ceiling Concentration, mg/kg	Pollutant Concentrations, mg/kg
Arsenic	75	41
Cadmium	85	39
Copper	4,300	1,500
Lead	840	300
Mercury	57	17
Molybdenum	75	NL
Nickel	420	420
Selenium	100	100
Zinc	7,500	2,800

NL = No established limit.

If the compost produced is Class B, it can be used at agronomic sites with no public contact, with additional site restrictions. Class A biosolids can be used in home gardens with public contact and no site restrictions. Consistent and predictable product quality is a key factor affecting the marketability of compost (30). Successful marketing depends on a consistent product quality. Stability is an important characteristic of a good quality compost. Stability is defined as the level of biological activity in the compost and is measured as oxygen uptake or carbon dioxide production. Oxygen uptake rates of 50 to 80 mg/L are indicative of a stable product with minimal potential for self-heating, malodor generation, or regrowth of pathogen populations. Stability is also indicated by temperature decline, ammonia concentrations, chemical oxygen demand (COD), number of insect eggs, change in odor, and change in redox potential (31).

Stable compost consumes little nitrogen and oxygen and generates little carbon dioxide. Unstable compost consumes nitrogen and oxygen and generates heat, carbon dioxide, and water vapor. Therefore, when unstable compost is applied to soil, it removes nitrogen from the soil, causing a nitrogen deficiency that can be detrimental to plant growth and survival. In addition, if not aerated and stored properly, unstable compost can emit nuisance odors (23, 32).

4. PROCESS DESCRIPTION

The basic composting process consists of the following steps (15):

1. The material to be composted must be porous, structurally stable, and capable of self-sustaining the decomposition reaction. If required, bulking agents for porosity and moisture control (for example, recycled compost, wood chips, etc.) or feed amendments for a source of limiting nutrients such as carbon (for example, sawdust, rice hulls, etc.) are added to the dewatered biosolids to provide a mixture suitable for composting.
2. Temperature in the range of 55° to 65°C (130° to 150°F) is required to ensure destruction of pathogenic organisms and provide the driving force for evaporation, which reduces the moisture content.

3. The compost is stored for extended periods after the primary composting operation to further stabilize the mixture at lower temperatures.
4. Additional air drying may be required if the cured compost is too wet for further processing.
5. When bulking agents are to be reused, a separation operation is required to remove the bulking agent from the compost at the end of the process.

The resulting product is generally cured for at least 30 days after active composting and before use. A properly operated facility produces a stable compost which can be easily handled and safely stored. Compost enhances soil properties, such as water holding capacity, nutrient availability, and texture. Because this process results in a usable material, an important and often overlooked part of any composting facility is product storage and marketing. Unlike disposal-oriented technologies, end users and markets for the product are seasonal with peak demand in the spring and fall. Therefore, provisions for storage of the final product until it is sold are necessary. In addition, product marketing efforts are essential to insure that end users understand the material, recognize its value, and are familiar with proper application techniques (18).

Composting represents the combined activity of a succession of mixed populations of bacteria, actinomycetes and other fungi associated with a diverse succession of environments. The principal factors which affect the microbiology of composting include (33):

1. Moisture
2. Temperature
3. pH
4. Nutrient concentration, and
5. Availability and concentration of oxygen.

4.1. Moisture

Decomposition of organic matter is dependent upon moisture. The lowest moisture content at which bacterial activity takes place is from 12 to 15 percent; however, less than 40 percent moisture may limit the rate of decomposition. The optimum moisture content is in the range of 50 to 60 percent. If the mixture is over 60 percent water, the proper structural integrity will not be obtained.

Dewatered municipal biosolids are usually too wet to satisfy optimum composting conditions. The moisture content can be reduced by blending the biosolids with a dry bulking material or a recycled product, and dewatering the biosolids to as great an extent as economically possible. The best approach for a particular site can be determined from a mass balance of the particular composting facility and by a site-specific economic analysis based on the mass balance results. Figure 16.1 illustrates the effect of the solids content of dewatered biosolids on the required mixing ratio of wood chips to biosolids by volume for one compost operation. The amount of wood chips needed for a 40 percent filter cake would be about one-fifth the amount required for a 20 percent solids cake. In addition to savings on wood chips, there would be a substantial reduction in material management costs and site sizes (34).

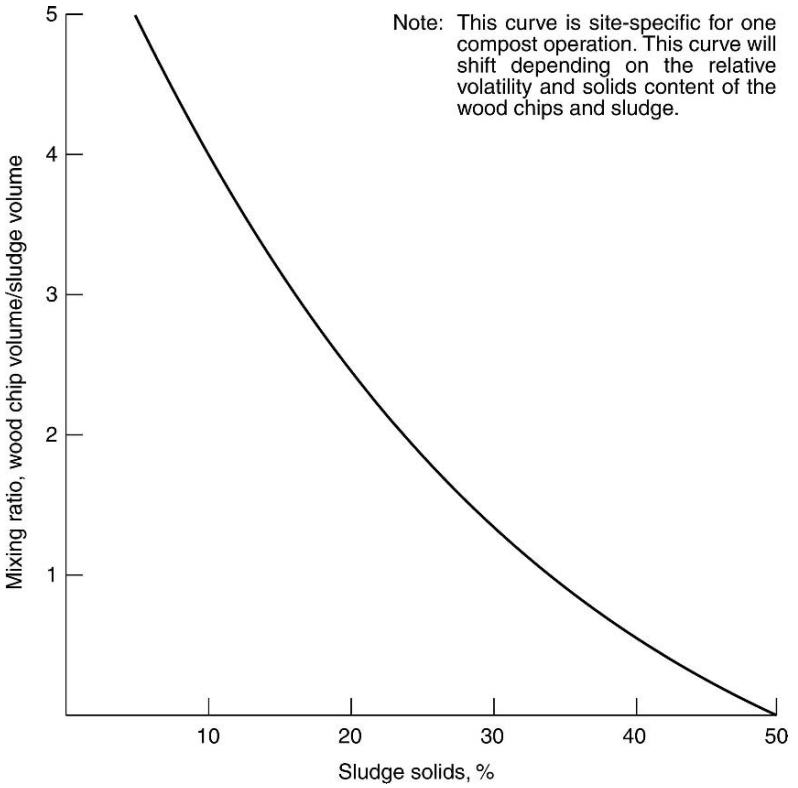


Fig. 16.1. Effect of solids content on the ratio of wood chips to biosolids by volume (15).

The U.S. Composting Council (35) lists the following materials as suitable for use as bulking agents:

1. Agricultural by-products, such as manure and bedding from various animals, animal mortalities, and crop residues.
2. Yard trimmings, including grass clippings, leaves, weeds, stumps, twigs, tree prunings, Christmas trees, and other vegetative matter from land clearing activities.
3. Food by-products, including damaged fruits and vegetables, coffee grounds, peanut hulls, egg shells, and fish residues.
4. Industrial by-products from wood processing, forestry, brewery and pharmaceutical operations. Paper goods, paper mill residues, and biodegradable packaging materials are also used.
5. Municipal solid waste.

4.2. Temperature

For most efficient operation, the temperature in the compost should range between 55° to 65°C (130° to 150°F) but not above 80°C (176°F). High temperatures are also required for the inactivation of human pathogens in the biosolids. The temperature distribution in a compost pile is affected by (15):

1. Moisture content
2. Aeration rates
3. Size and shape of pile
4. Atmospheric conditions, and
5. Nutrients

For example, temperature elevation will be less for a given quantity of heat released if excessive moisture is present, as heat will be carried off by evaporation. On the other hand, low moisture content will decrease the rate of microbial activity and thus reduce the rate of heat evolution.

4.3. pH

The optimum pH range for growth of most bacteria is between 6 and 7.5 and for fungi between 5.5 and 8.0 (36). The pH varies throughout the pile, and throughout the composting operation, but it is essentially self regulating. A high initial pH resulting from the use of lime for dewatering will solubilize nitrogen in the compost and contribute to the loss of nitrogen by ammonia volatilization. It is difficult to alter the pH in the pile for optimum biological growth, and this has not been found to be an effective operation control.

4.4. Nutrient Concentration

Both carbon and nitrogen are required as energy sources for organism growth. Thirty parts by weight of carbon (C) are used by microorganisms for each part of nitrogen (N); a C/N ratio of 30 is, therefore, most desirable for efficient composting, and C/N ratios between 25 and 35 provide the best conditions (1). The carbon considered in this ratio is biodegradable carbon. Lower C/N ratios increase the loss of nitrogen by volatilization as ammonia and higher values lead to progressively longer composting times as nitrogen becomes growth-rate limiting (33). No other macro-nutrients or trace nutrients have been found to be rate limiting in composting municipal wastewater biosolids.

4.5. Oxygen Supply

Optimum oxygen concentrations in a composting mass are between 5 and 15 percent by volume (37). Increasing the oxygen concentration beyond 15 percent by air addition will result in a temperature decrease because of the greater air flow. Although oxygen concentrations as low as 0.5 percent have been observed inside windrows without anaerobic symptoms, at least 5 percent oxygen is generally required for aerobic conditions (33).

5. DESIGN CRITERIA AND PROCEDURES

The basic criteria for successful composting are that the material to be composted be porous and structurally stable and contain sufficient degradable material so that the degradation reaction is self-sustaining (that is, heat released by oxidation of volatile material is sufficient to raise the mixture to reaction temperature and to bring it to required dryness). In this section, a procedure to meet these criteria of porosity, structural stability, and sufficient biodegradability will be discussed. An equally important design consideration is flexibility.

Table 16.3
Monitoring program for a municipal wastewater biosolids composting facility (15)

Activity/time	Component	Analysis	Frequency
Before composting	Sludge and bulking material	Heavy metals and PCB's	Monthly
During composting	Aerated pile or windrows	Acceptable time, temperature, dissolved oxygen relationships, that is, 131°F (55°C) and 5 to 15 percent oxygen content for 3 to 5 days.	Temperature and oxygen content measurements taken at least 6 days during first 2 weeks. (Additional measurements sometimes required to get true average).
After composting	Compost (prior to marketing)	Certain selected indicator heavy metals and pathogens.	Monthly or bimonthly depending on use of compost.
Site monitoring during entire operation	Personnel	Physical examination prior to employment and periodically thereafter.	Annually
		Protective equipment and clothing as needed.	Continuously
	Odors	Odor strength	Continuously, but especially during wet periods with temperature inversions and little to no wind.
		Odor filter pile effectiveness.	Continuously
		Log of odor complaints.	Continuously
	Dust	Assessment of particulate concentrations.	Continuously but especially during dry period under windy conditions.
	Leachate and runoff	BOD and suspended solids.	Monthly, downwind at locations critical to public health concerns.
	Airborne spores	Numbers generated and transported.	Monthly
	Micrometeorological	Temperature at 5 ft (1.5 m) and 25 ft (7.6 m).	Continuously
		Wind speed	Continuously
		Wind direction	Continuously

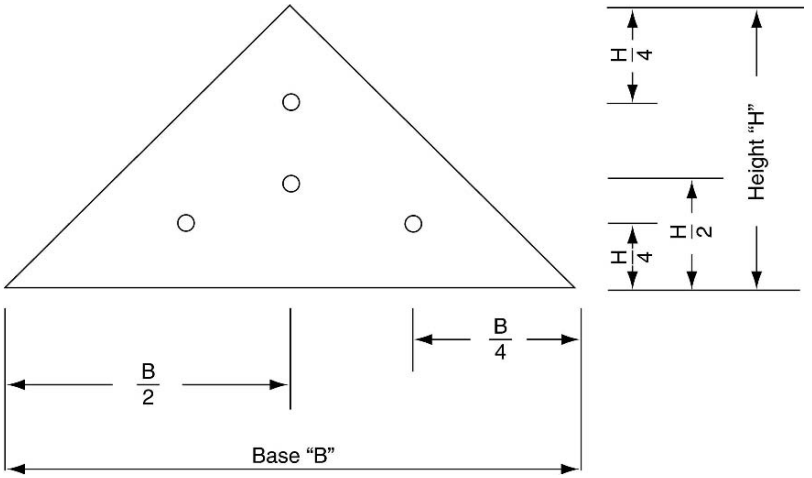


Fig. 16.2. Locations for temperature & oxygen monitoring at one end of a windrow or individual aerated pile (15).

A compost operation must be able to operate continuously even with changes in biosolids content and volume. Changes in bulking agent supply and equipment failure must also be anticipated, and the design must be flexible to deal with these changes (15).

To obtain minimal assurance that the composting activity is proceeding properly, the temperature and oxygen content within the pile are constantly monitored. Equipment required to conduct this monitoring includes a portable, 0 to 25 percent, dry-gas oxygen analyzer which is used to measure the oxygen content; a probe-thermistor-type temperature indicator, with at least a 6-ft probe and scale reading from 32° to 212°F (0° to 100°C) is also needed. Additionally, monitoring of heavy metals, pathogens, and environmental parameters such as air and water quality ensures safe and acceptable compost and composting operation. Oxygen respirometry to assess stability and maturity of composted biosolids is recommended (38). A comprehensive monitoring program is outlined in Table 16.3. Four locations for temperature and oxygen measurements at both ends of each pile are shown on Figure 16.2.

Haug and Haug (39) have shown the compost reaction is self-sustaining when the ratio W is ≤ 10:

where:

W = mass of water in compost mixture/mass of organics degraded by composting

In windrow and mechanical composting, porosity and structural stability are provided when the biosolids are mixed with recycled compost product or bulking agent to obtain solids concentration of approximately 40 to 60%. With aerated pile composting, a bulking agent such as wood chips is used to provide porosity and structural stability. When the composting process is complete, the bulking agents are generally screened out of the compost and recycled back to the mix point for reuse. The fine portion of the bulking agent is usually retained with the compost product because it passes through the screen with the finished compost. Fresh bulking agent must be added at the mix point to compensate for this material loss.

Mixture degradability can be adjusted by the addition of materials that contain high concentrations of degradable organic material. These materials are usually dry and reduce the ratio W by increasing the volatile fraction and decreasing the moisture fraction of the mixture.

Figure 16.3 shows a generalized mass balance diagram for the compost process. The recycle stream could consist of finished compost only (typical for windrow and mechanical methods), bulking agent only (typical for aerated pile methods) or a combination of bulking agent and finished compost. Amendment may also be added with bulking agent. The exact quantities of the various streams are dependent on the mass balance equations (1) and (2) derived from Figure 16.3 and the type of composting process utilized.

A set of equations can be developed from an analysis of the mass balance diagram. Two general equations have been arranged that apply to all composting methods. Equation (1) is used to determine the recycled compost or wood chip quantity and Equation (2) is used to determine the ratio W (39):

$$X_R = X_C(S_M - S_C) + X_A(S_M - S_A) + X_B(S_M - S_B)/(S_R - S_M) \quad (1)$$

$$W = [X_C(l - S_C) + X_A(l - S_A) + X_B(l - S_B) + X_R(l - S_R)]/$$
$$[X_C S_C V_C k_C + X_A S_A V_A k_A + X_B S_B V_B k_B + X_R S_R V_R k_R] \quad (2)$$

5.1. Compost Processes with no External Bulking Agent

To design a compost facility employing no external bulking agent, the parameters X_C, S_C, V_C, k_C, S_R, V_R, k_R, and S_M must be determined analytically, assumed, or calculated. The wet weight of recycled compost (X_R) is calculated, assuming no amendment or external bulking agent addition ($X_A = X_B = O$), to provide a desired solids content of the mixture (S_M) in the 0.40 to 0.50 range:

$$X_R = X_C(S_M - S_C)/(S_R - S_M) \quad (3)$$

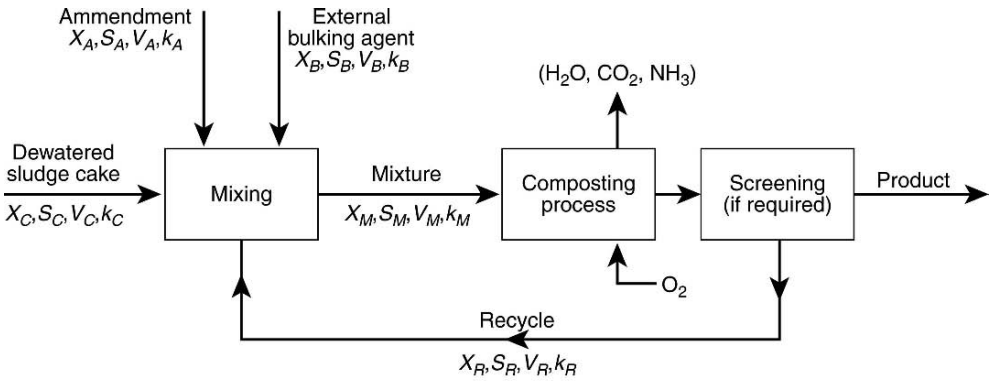
Once X_R is determined for these conditions, the ratio W is calculated:

$$W = [X_C(l - S_C) + X_R(l - S_R)]/[X_C S_C V_C k_C + X_R S_R V_R k_R] \quad (4)$$

If the ratio W is less than ten, the compost mixture has sufficient energy available for temperature elevation and water evaporation. The ratio number of ten is not absolute because climatic conditions affect the thermodynamic energy requirements. In a hot, arid climate, W may be higher because evaporation of water from the compost mass is increased by a high humidity driving force and higher initial pile temperatures. In a cold climate, more biological energy is required to heat the pile to normal operating temperatures and thus W may have to be as low as seven to ten (39).

The ratio W can be reduced by adding amendment. The parameters S_A, V_A, and k_A are known. The amendment dry weight is assumed and new recycle compost mass (X_R) is calculated:

$$X_R = [X_C(S_M - S_C) + X_A(S_M - S_A)]/(S_R - S_M) \quad (5)$$



Note: RECYCLE is defined as finished compost for the windrow and mechanical systems and as recycled wood chips for the aerated pile system.

The exact value for these parameters must be determined from samples of the sludge, external bulking agent, amendment, and estimated for the recycle values unless otherwise known.

Process Variables and Range of Average Values (in Parenthesis)

X_C = Total wet weight of sludge cake produced/d.

X_A = Total wet weight of amendment/d.

X_R = Total wet weight of recycle/d.

X_B = Total wet weight of external bulking agent/d.

X_M = Total wet weight of mixture/d.

S_C = Fractional solids content of sludge cake (0.20–0.55).

S_A = Fractional solids content of amendment (0.50–0.95).

S_R = Fractional solids content of recycle (0.60–0.75).

S_B = Fractional solids content of external bulking agent (0.50–0.85).

S_M = Fractional solids content of mixture (0.40–0.50).

V_C = Volatile solids content of sludge cake, fraction of dry solids (0.40–0.60)–Digested; (0.60–0.80)–Raw.

V_A = Volatile solids content of amendment, fraction of dry solids (0.80–0.95).

V_R = Volatile solids content of recycle fraction of dry solids (0–0.90).

V_B = Volatile solids content of recycle external bulking agent, fraction of dry solids (0.55–0.90).

V_M = Volatile solids content of mixture, fraction of dry solids (0.40–0.80).

k_C = Fraction of sludge cake volatile solids degradable under composting conditions (0.33–0.56).

k_A = Fraction of amendment volatile solids degradable under composting conditions (0.40–0.60).

k_R = Fraction of recycle volatile solids degradable under composting conditions (0–0.20).

k_B = Fraction of external bulking agent volatile solids degradable under composting conditions (0–0.40).

k_M = Fraction of mixture volatile solids degradable under composting conditions (0.20–0.60).

Fig. 16.3. Biosolids composting mass balance diagram (15).

The ratio W is also recalculated:

$$W = [X_C(1 - S_C) + X_R(1 - S_R) + X_A(1 - S_A)]/$$
$$[X_C S_C V_C k_C + X_R S_R V_R k_R + X_A S_A V_A k_A] \quad (6)$$

If W is still not below ten, the quantity of amendment is increased and X_R and W are recalculated until the W requirement is satisfied.

If these guidelines are followed, a mixture with sufficient energy to compost will be produced. The actual values for the process parameters are site-specific and the most economical design is dependent on accurate information about the composting characteristics that affect the mass and thermodynamic balance.

5.2. Compost Processes Using External Bulking Agent

Design criteria for processes using external bulking agent are similar to those just described except that the recycle rate is calculated in a different manner. In the former process, the ratio of total bulking agent to biosolids is specified without regard to the mixture's moisture content, since it is not as important as the structural integrity of the pile. The recycle rate, X_R, and makeup supply are calculated using Equations (7) and (8).

$$X_R = (1 - f_2) f_1 X_C \quad (7)$$

$$X_B = f_1 X_C - X_R \quad (8)$$

where f_1 is defined as the ratio of external bulking agent (recycle and makeup) to biosolids:

$$f_1 = X_R + X_B / X_C \quad (9)$$

and f_2 represents the fraction of total external bulking agent lost from the process by volatilization or because it remains with the finished compost:

$$f_2 = X_B/[X_B + X_R] \quad (10)$$

The values for f_1 and f_2 must be assumed based on operating experience at an existing facility. The range of values for f_1 is 0.75 to 1.25, and for f_2 is 0.20 to 0.40. Once these values are chosen, the amount of recycled bulking agent (X_R) and new external bulking agent (X_B) can be calculated using Equations (7) and (8).

The value of the ratio W is then calculated using Equation (2), indicating no amendment is used ($X_A = 0$). If W is less than or equal to ten, then the mixture has sufficient energy to compost. If W is greater than ten, two options for reducing the ratio are possible. More external bulking agent can be used (that is, f_1 is increased). If the bulking agent is more volatile than the biosolids, W should be reduced. The recycle and makeup quantities of bulking agent must be recalculated and W determined again. If the bulking agent is of low volatile fraction, this approach will not work because W will be reduced only slightly. In this case, amendment must be added.

For any amount of amendment addition, the ratio W can again be calculated using Equation (2). Increasing the amount of amendment until W is below ten will result in the proper compost energy balance.

Table 16.4
Densities of compost and bulking agents (15)

Material	Density, lb/yd^3
Digested sludge	1,500 to 1,700
Raw sludge	1,300 to 1,700
New wood chips	445 to 560
Recycled wood chips	590 to 620
Finished compost	930 to 1,040

The operation at Bangor, Maine, successfully composts biosolids by the aerated pile method in winter months. No amendment is used, and the ratio of external bulking agent (bark) to biosolids by volume is 2.5:1. The value for W ranges from seven to ten at this operation (39).

The best means to determining the quantities of external bulking agent and amendment used will be a careful economic analysis of the process and accurate estimation of the process variables. Table 16.4 lists the average density for various compost materials as experienced at various compost facilities.

6. WINDROW PROCESS

In the United States, the windrow and aerated static pile processes have been used almost exclusively for composting dewatered municipal wastewater biosolids. The basic steps to be followed in these two processes are similar, but the processing technology for the composting stage differs appreciably. In the windrow method, oxygen is drawn into the pile by natural convection and turning, whereas in the static pile method, aeration is induced by forced air circulation.

The windrow process is normally conducted in uncovered areas and relies on natural ventilation with frequent mechanical mixing of the piles to maintain aerobic conditions. In areas of significant rainfall, it may be desirable for operational reasons to provide a roofed structure to cover the windrows for composting biosolids. The largest operating windrow process in the United States is located at the Joint Water Pollution Control Plant of the County Sanitation Districts of Los Angeles County in Carson, California (15).

In the process biosolids are converted to a relatively stable organic residue, reduced in volume by 20 to 50 percent. The residue loses its original identity with respect to appearance, odor and structure. The end product has earthy characteristics; pathogens, weed seeds and insect larvae are destroyed (40).

6.1. Methodology and Design

In the windrow composting process, the mixture to be composted is stacked in long parallel rows or windrows. The cross section of the windrows may be trapezoidal or triangular, depending largely on the characteristics of the mobile equipment used for mixing and turning the piles. The width of a typical windrow is 15 ft (4.5 m) and the height is 3 to 7 ft (1 to 2 m).

Based on processing biosolids containing 20 percent solids, land requirements for the windrow process are greater than for the aerated pile process. It has been estimated that an extra 25 percent land area is needed for the windrow process based on windrows 5 ft (1.5 m) high and 7 ft (2 m) wide with a two-week composting period (41). Even more land would be necessary for the longer composting time experienced in the Los Angeles operations.

The mixing of a bulking agent with the wet biosolids cake has enabled the windrow process to be used for composting digested dewatered biosolids. Bulking agents may include the recycled composted biosolids itself or external agents such as wood chips, sawdust, straw, rice hulls, or licorice root. The quantity of bulking agent is adjusted to obtain mixture solids content of 40 to 50%. The use of a bulking agent also increases the structural integrity of the mixture and thus, its ability to maintain a properly shaped windrow. Porosity of the mixed material is greatly improved, which in turn improves the aeration characteristics. External bulking agents can also provide a source of carbon for the composting process. The carbon to nitrogen (C/N) ratio of digested activated sludge is in the range of 9 to 15:1. If wood chips are used as the bulking agent, the C/N ratio will be raised to approximately 20 to 30:1 in the composting mixture (15).

Convective air movement within windrows is essential for providing oxygen for the microorganisms. The aerobic reaction provides heat for warming the windrows. This causes the air to rise, producing a natural chimney effect. The rate of air exchange can be regulated by controlling the porosity and size of the windrow (3). The turning of the windrow also introduces oxygen to the microorganisms. This method of aeration can be expensive if used excessively to obtain high oxygen concentrations and may reduce the temperature within the windrow. A number of turning devices are available, including: (a) drums and belts powered by agricultural equipment and pushed or pulled through the composting pile; and (b) self-propelled models that straddle the composting pile (1).

As a result of the biological decay process, temperatures in the central portion of the windrow reach as high as 150°F (65°C). Operating temperatures of about 140°F (60°C) may be maintained in the central portion of the windrow for as long as ten days. Temperatures in the outer layers are considerably cooler and may approach atmospheric conditions. During wet periods and winter conditions, maximum temperatures may only be 130° to 140°F (55° to 60°C). A high temperature maintained throughout the pile for a sufficient period of time is important to the control of pathogens. A satisfactory degree of stabilization is indicated by a decline in temperature, usually to about 113° to 122°F (45° to 50°C). These variations in temperature are illustrated in Figure 16.4.

Large-scale, 270 dry ton/d (243 t/d) processing of digested primary biosolids (23 percent solids) using the windrow process, with recycled composted biosolids as the bulking agent, has proven a viable method of biosolids stabilization by the Los Angeles County Sanitation Districts. Successful operation of the windrow process using bulking agents such as wood chips and sawdust with digested primary and secondary biosolids has also been achieved at Beltsville. This process has not proven suitable for composting unstabilized primary or secondary biosolids. At Beltsville during early tests with windrows, undigested primary and waste activated sludge biosolids were found to produce offensive odors (4). Also, composting of digested biosolids did not kill all seeds, and these were present in the final product.

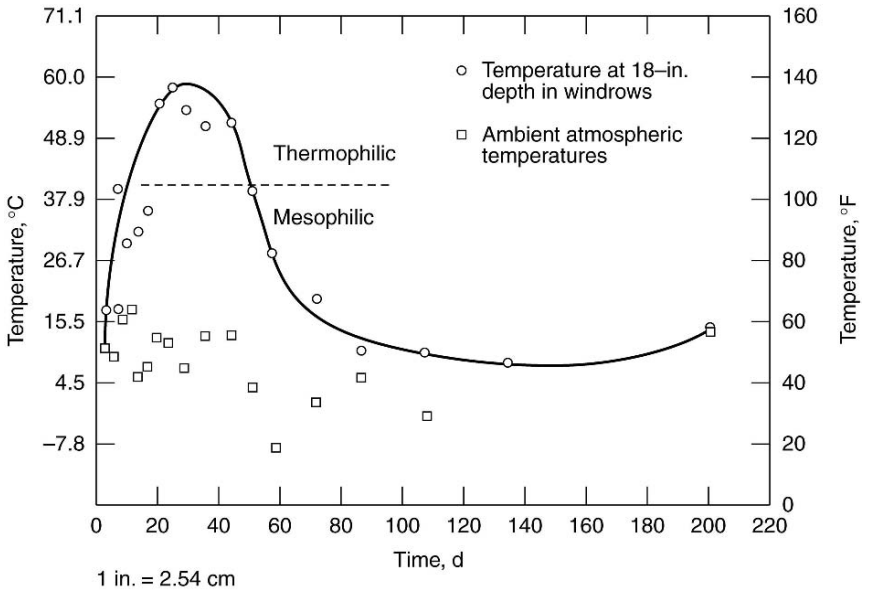


Fig. 16.4. Temperature profile of a typical compost windrow (15).

The Los Angeles County Sanitation Districts are currently composting digested, centrifuged primary biosolids (23 percent solids) in windrows mixed with recycled composted biosolids (60 percent solids) in a 1:2.2 ratio (dry weight). A compost mixing machine is used to turn the mixture. Recycled compost is added to the biosolids before the windrow is constructed. Each windrow must be turned two or three times a day for the first five days to mix the material completely, minimize odors, and ensure sufficient oxygen transfer. The biosolids are then turned once a day for about 30 days, depending on weather conditions.

Large, portable, heavy materials handling equipment is required for the windrow system. The Los Angeles operation requires four windrow mixing-turning machines capable of turning 3,400 tons/h (3,084 t/h) of a density of 1,890 lb/yd^3 (1,120 kg/m^3). This is equivalent to a volume capacity of 3,600 yd^3/h (2,752 m^3/h). Three machines operate continuously for two shifts a day. A fourth machine is required to provide backup whenever any of the others is being repaired. In case of rain all four machines must operate continuously.

Sawdust, shredded paper, and wood chips were the external bulking agents used in the Beltsville windrow tests. Only shredded paper was found to be unsatisfactory (3). The windrow area at Beltsville was paved with 18 inches (0.46 m) of crushed stone to support heavy equipment and the windrow composter. The area was later paved with asphalt and then with concrete to assure positive leachate collection and to eliminate rock pickup from the collection equipment and damage to the screening equipment. To start the windrow, a layer of wood chips 15 inches (0.38 m) deep and 15 ft (4.5 m) wide was placed on the paved area. Biosolids (20 to 25 percent solids) were distributed to the chips at a 1:3 volume ratio. The compost machine then mixed the biosolids and chips. After several turnings, the two

materials were thoroughly mixed. The windrow was turned five times a week, flattened after two weeks to a 12-inch (0.30 m) layer and harrowed for further drying, generally to greater than 65 percent solids. The material was then removed from the windrow area and stockpiled for an additional 30 days for curing purposes. Curing was required to improve compost quality and to further control pathogens. After curing, the composted mixture was distributed to local government agencies as screened or unscreened material. Wood chips separated during the screening operation were recycled and reused as bulking agent. The use of a bulking agent may substantially increase the cost of the composting process unless the bulking agent is itself a waste material (8). At Beltsville, a fresh supply of wood chips was required to make up for the estimated 25 to 30 percent lost in the composting process. Some of the bulking agent was consumed in the biological oxidation processes during composting, and a large portion was lost in the screening process.

6.2. Energy Requirements

Thermodynamic considerations in the composting of biosolids are discussed in an article by Haug & Haug (39). As indicated previously, the reaction is self-sustaining when the ratio W is less than ten. Over 80 percent of the heat released by the biological reaction is used to evaporate moisture associated with the biosolids.

In the windrow process, the only external energy requirements are gasoline for transportation, diesel fuel for operation of composting machines, and electricity for leachate treatment and site services, including lighting. In the Beltsville windrow tests, which used wood chips as a bulking agent, the following operating requirements per dry ton/d (0.9 T/d) for a 10 to 50 dry ton/d (9 to 45 T/d) operation been estimated (41):

(a) Labor: 1.8 to 3.0 h
(b) Gasoline: 1.1 gall (4.5 L)
(c) Diesel Fuel: 3.3 to 4.0 gall (13.5 to 16.5 L)
(d) Electricity: 3.0 to 8.0 kwh (12 to 32 MJ)

Where finished compost is used as the bulking agent, and increased windrow turning frequency is practiced, higher diesel fuel consumption should be expected.

6.3. Public Health and Environmental Impacts

Numerous studies have indicated that a community's wastewater contains organisms which reflect the local prevalent endemic diseases (42). The pathogens borne by wastewater are not entirely inactivated during conventional biosolids digestion and drying techniques and may persist in the soil for extended periods of time. Figure 16.5 shows this time-temperature-destruction relationship of pathogens for windrows (43, 44).

Intensive studies conducted by the Los Angeles County Sanitation Districts indicate that total coliform and Salmonella concentrations are rapidly reduced in the first ten days of composting in the interior of windrows. For interior samples, final compost coliform concentrations of less than one per gram have been attained, but higher values for exterior samples have been measured consistently. Very low levels of virus, parasitic ova, and Salmonella have been assayed in the majority of final compost samples.

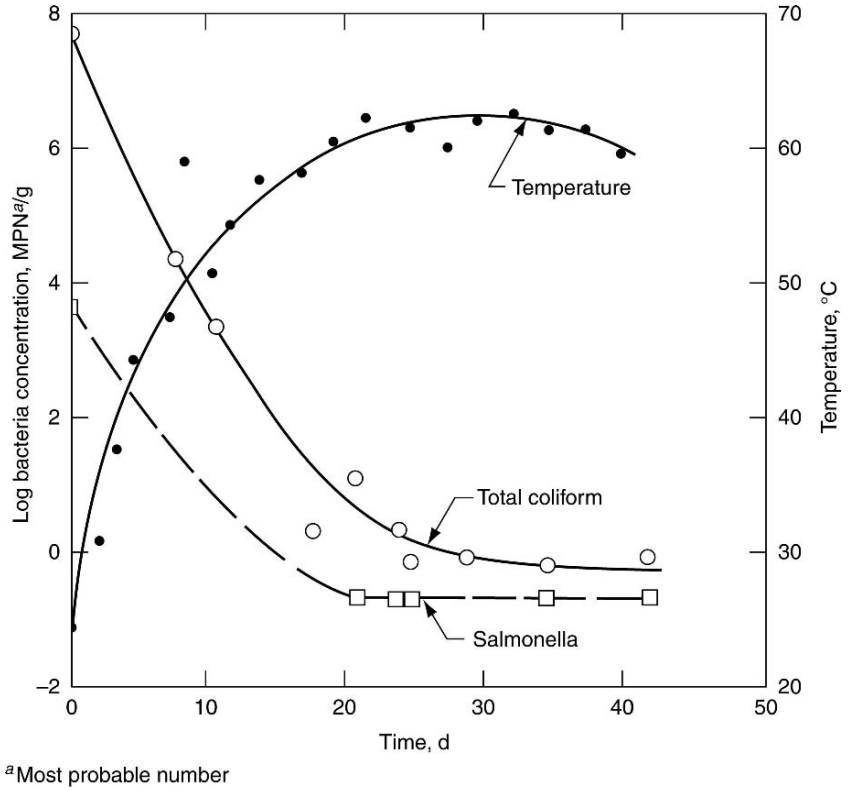


Fig. 16.5. Destruction of pathogenic organisms as a function of time and temperature during the composting of undigested biosolids by the windrow method (15).

Recycling large quantities of finished compost as bulking agent provides good odor control for digested biosolids, as long as process upsets are kept under control. Interruption of regular turning of the biosolids may cause odor problems, since compost windrows quickly become anaerobic under these circumstances. Unpleasant odors may also be generated during periods of high rainfall, as well as by poor mixture control and inefficient mixing. In dry and windy areas, wetting of the compost windrows should be practiced to prevent excessive dust generation.

A drainage and collection system is required for stormwater runoff from the site because the contaminated water requires treatment. The runoff may be returned to the wastewater treatment plant. At Beltsville, a wooded area adjacent to the site was spray irrigated (3).

Workers at a compost site should avoid inhaling dust. Respiratory protection, such as breathing masks, should be worn in dusty areas, and the area should be sprinkled with water during dry periods. Although recent experiments have shown high concentrations of the fungus *Aspergillus fumigatus*, a secondary pathogen, to be airborne at biosolids composting sites, preliminary data indicate that these higher spore levels are generally restricted to the

immediate composting area and should not pose a significant health threat to surrounding residential, commercial, or industrial areas (45). However, individuals with a history of lung ailments should not work in composting operations. Research is continuing on potential health effects of exposure to the fungus *A. fumigatus* (46 to 50).

7. AERATED STATIC PILE PROCESS

7.1. Process Description

An aerated static pile system was developed in order to eliminate many of the land requirements and other problems associated with the windrow composting process and to allow composting of raw biosolids. A diagram of an aerated pile for composting biosolids is shown in Figure 16.6.

Wastewater biosolids are converted to compost in approximately eight weeks in a four-step process (40)

(a) *Preparation* – Biosolids are mixed with a bulking material such as wood chips or leaves, in order to facilitate handling, to provide the necessary structure and porosity for aeration, and to lower the moisture content of the biomass to 60 percent or less. Following mixing, the aerated pile is constructed and positioned over porous pipe through which air is drawn. The pile is covered for insulation.

(b) *Digestion* – The aerated pile undergoes decomposition by thermophilic organisms, whose activity generates a concomitant elevation in temperature to 60°C (140°F) or more. Aerobic composting conditions are maintained by drawing air through the pile at a predetermined rate. The effluent air stream is conducted into a small pile of screened, cured compost where odorous gases are effectively absorbed. After about 21 days the composting rates and temperatures decline, and the pile is taken down, the plastic pipe is discarded, and the compost is either dried or cured depending upon weather conditions.

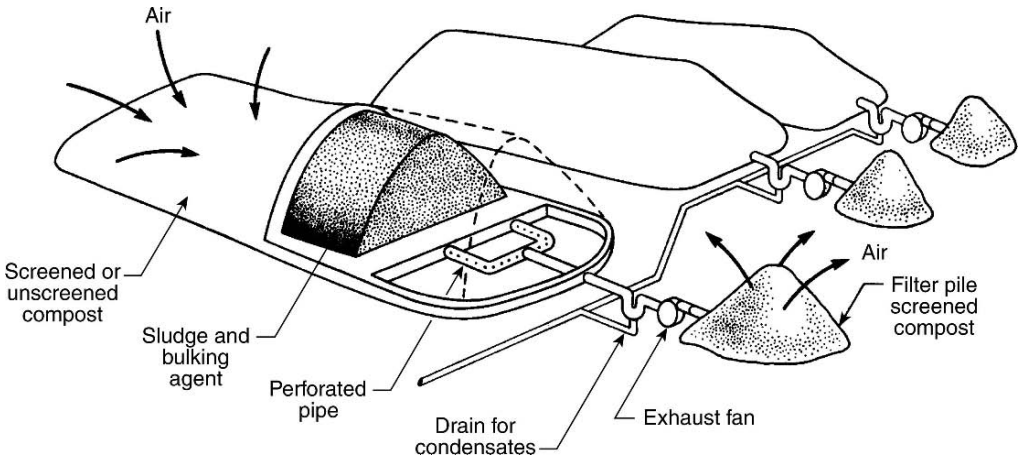


Fig. 16.6. Configuration of individual aerated piles (15).

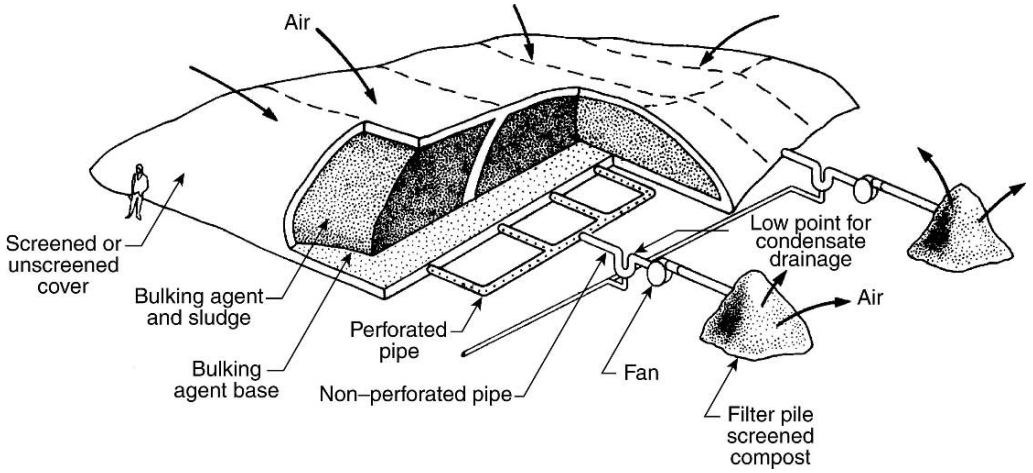


Fig. 16.7. Aeration pipe set-up for individual aerated pile (15).

(c) *Drying and Screening* – Drying to 40 to 45 percent moisture facilitates clean separation of compost from wood chips. The unscreened compost is spread out with a front end loader to a depth of 12 inches. Periodically a tractor-drawn harrow is employed to facilitate drying. Screening is performed with a rotary screen. The chips are recycled.

(d) *Curing* – The compost is stored in piles for about 30 days to assure no offensive odors remain and to complete stabilization. The compost is then ready for utilization as a low grade fertilizer, a soil amendment, or for land reclamation.

The forced air method provides for more flexible operation and more precise control of oxygen and temperature conditions in the pile than would be obtained with a windrow system (51). Since composting times tend to be slightly shorter and anaerobic conditions can be more readily prevented, the risk of odors is reduced. Two distinct aerated static pile methods have been developed, the individual aerated pile and the extended aerated pile.

7.2. Individual Aerated Piles

An individual aerated pile may be constructed in a manner similar to the Beltsville method, in which loop of perforated plastic pipe, 4 to 6 inches (10 to 15 cm) in diameter is placed on the composting pad, oriented longitudinally, and centered under the ridge of the pile under construction. In order to avoid short circuiting of air, the perforated pipe terminated at least 8 to 10 ft (2 to 3 m) inside the ends of the pile. A non-perforated pipe that extends beyond the pile base is used to connect the loop of perforated pipe to the blower (See Figure 16.7).

A 6- to 8-inch (15 to 20 cm) layer of bulking agent is placed over both the pipes and the area to be covered by the pile. This base facilitates the movement and even the distribution of air during composting and absorbs excessive moisture that may otherwise condense and drain from the pile (42).

At Beltsville a mixer or front-end loader is used to mix one volume of biosolids cake containing 22 percent solids and two volumes of bulking agent. The resulting mixture contains

40 percent solids and is placed loosely upon the prepared base by the front-end loader to form a pile with a triangular cross section 15 ft (4.6 m) wide by 7.5 ft (2.3 m) high.

The pile is then completely covered with a 12-inch (0.3 m) layer of cured, screened compost or an 18-inch (0.4 m) layer of unscreened compost. This outer blanket of compost provides insulation and prevents escape of odors during composting. Unstabilized biosolids can generate odors during dumping and initial pile construction. Conditioning with lime during dewatering will minimize this, however. The non-perforated pipe is connected to a 1/3 hp (0.25 kW), 335 ft^3/min (158 L/s) blower that is controlled by a timer (52). Aerobic composting conditions are maintained if air is intermittently drawn through the pile. The timing sequence for the blower is 5 minutes on and 15 minutes off for a 56-ft (17 m) long pile containing up to 80 wet tons (73 t) of biosolids. If the aeration rate is too high or the blower remains on too long, the pile will cool, and the thermophilic process will be inhibited (33).

The effluent air from the compost pile is conducted into a small, cone-shaped filter pile of cured, screened compost approximately 4 ft (1.2 m) high and 8 ft (2.5 m) in diameter where malodorous gases are absorbed. The odor retention capacity of these piles is inhibited if their moisture content is greater than 50 percent. The odor filter pile should contain one cubic yd (0.76 m^3) of screened compost for each four dry tons (3.6 T) of biosolids in the compost pile. Filter piles are sometimes constructed with a 4-inch (10 cm) base layer of wood chips to prevent high back pressures on the blower.

Land area requirements are estimated at one acre per 3 to 5 dry tons (1.0 ha/6.7 to 11.2 T) of biosolids treated. The lower figure includes space for runoff collection, administration, parking, and general storage. The actual composting area (mixing area, aerated piles, screening area, drying area, and storage area) is estimated to be one acre per 5 dry tons (1.0 ha/11.2 T) of biosolids (42).

7.3. Extended Aerated Piles

To make more effective use of available space, another static pile configuration called the extended aerated pile has been developed. An initial pile is constructed with a triangular cross section utilizing one day's biosolids production. Only one side and the ends of this pile are blanketed with cured, screened compost. The remaining side is dusted with only about an inch (0.5 cm) of compost for overnight odor control. The next day, additional aeration pipe is placed on the pad parallel to the dusted side of the initial pile. The pile bed is extended by covering the additional pipe with more bulking agent and biosolids-bulking agent mixture so as to form a continuous or extended pile. This process is repeated daily for 28 days. The first section is removed after 21 days. After seven sections are removed in sequence, there is sufficient space for operating the equipment so that a new extended pile can be started. Figure 16.8 shows such a system. The area requirement of an extended pile system is about 50 percent less than that for individual piles. The amount of recycled bulking agent required for covering the pile and bulking agent used in the construction of the base is also reduced by about 50 percent.

The aerated pile system has proven effective on a full-scale basis at Beltsville, Maryland; Bangor, Maine; Durham, New Hampshire; Detroit, Michigan; and Windsor, Ontario. After start-up, mean temperature in aerated piles is 176°F (70°C); and after stable conditions are

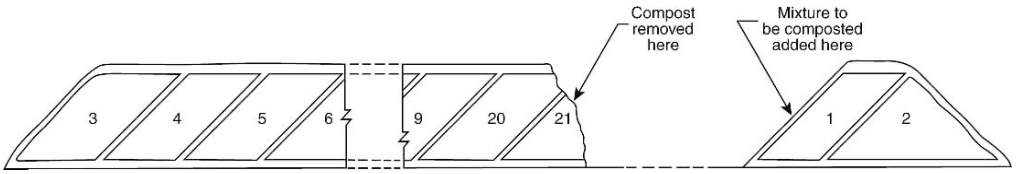


Fig. 16.8. Configuration of extended aerated pile (15).

achieved, minimum temperature is usually 130°F (55°C). When the piles are constructed properly, neither excessive rainfall nor low ambient temperature adversely affects the composting process (52).

The applicability of this system for the treatment of undigested biosolids provides it with a significant advantage over the windrow method. Other advantages are superior odor control, greater inactivation of pathogenic organisms, and use of less site area. The aerated pile technique exposes all biosolids to more uniform temperature. Capital costs are also lower for the aerated pile system, but operating costs tend to be higher because of the cost of the bulking agent. In experiments at Los Angeles County, it has been found necessary to follow this technique by windrow composting for 2 to 3 days to dry off the moisture. At other locations, the air flow is reversed without disruption of the pile as another means to reducing moisture content.

7.4. Oxygen Supply

Centrifugal fans efficiently provide the necessary pressure to move air through the compost and odor filter piles. Variation in the blower pressure is a necessity for optimum conditions and a site-specific operating parameter. The oxygen concentration in the pile should be maintained between 5 and 15 percent; this can be achieved with an aeration rate of about 500 $ft^3/h/t$ (15.6 $m^3/h/T$) dry biosolids. If the pile cools at this air rate, the air flow must be reduced. Aeration cycles of 20 to 30 min with the fan operating 1/10 to 1/2 of the cycle have proven satisfactory (42). While the fan is not operating, the natural convective chimney effect, typical of windrows, takes place. In the absence of forced aeration, this effect causes warming of the outer edges, destroying pathogens more effectively.

Moist air drawn through the pile condenses in the slightly cooler sections. When enough condensate accumulates, it will drain from the pile and leach material from the biosolids. Condensed moisture which collects in the aeration pipes is removed by a water trap. This material must be collected and treated along with the contaminated rainfall runoff from the site, because it can become a source of odors if allowed to accumulate in puddles around the piles. Data is not available on combined leachate and condensate water characteristics; the quantity may, however, vary from 6 to 20 gal/d (22 to 75 L/d) per pile containing 50 yd^3 (38 m^3) of biosolids during dry weather (53).

7.5. Bulking Agent

While bulking agents are in the aerated pile composting system, they serve primarily to maintain the structural integrity and porosity of the pile. The quantity of external bulking

agent required is determined by the need for structural support and porosity. The requirements for moisture control are not as critical as adequate porosity; thus, biosolids moisture can vary considerably as long as sufficient bulking agent is added to assure adequate porosity. The design factors discussed for windrows do not apply here (39).

Wood chips and other bulking agents also increase the volatile solids content of the composting mixture; volatility of new and recycled wood chips has been reported as 90 and 86 percent, respectively (41). The actual contribution of the wood chips to the compost mixture is limited because their composting rate is slower.

When wood chips are mixed with unstabilized biosolids an average volatility of about 75 percent results; this is well in excess of the 40 to 50 percent volatility achieved in the mixture of digested biosolids and recycled compost. Volatility content is therefore not a limiting factor in aerated pile composting of unstabilized biosolids, as it can be in the digested biosolids windrow system.

7.6. Energy Requirements

Energy costs for aerated pile composting are a small portion of the overall operating costs. The bulk of the overall energy requirement of the process is provided by the volatile solids in the composting mixture. A range of operating requirements per dry ton/d (0.9 T/d) for a 10 to 50 dry ton/d (9 to 45 T/d) operation (20% biosolids) is listed below (41):

(a) Labor: 1.5 to 2.8 h
(b) Wood Chips: 2 to 8 yd^3 (2.1 m^3)
(c) Gasoline: 1.1 gal (4.1 L)
(d) Diesel Fuel: 2.7 to 3.5 gal (10.2 to 13.2 L)
(e) Electricity: 7.5 to 17.5 kwh (29.7 to 69.3 MJ)

7.7. Public Health and Environmental Impacts

Extensive studies have been made on the destruction of pathogens in aerated piles (54). Although Salmonella, fecal coliforms, and total coliforms initially increased in numbers, they were reduced to essentially undetectable levels by the tenth day of composting. Studies using "F" bacteriophage and virus as an indicator showed that the virus was essentially destroyed by the thirteenth day. However, survival of the virus did occur for some time in the blanket-compost interface where lower temperatures prevailed. Storage in a curing pile for 30 days will complete the destruction of viruses or reduce the numbers to an extremely low level (42. Studies have shown that the composting process in an aerated pile is essentially unaffected by low ambient temperatures or rainfall, which makes this system particularly well suited to operation under difficult climatic conditions (55). Figure 16.9 shows the time-temperature-destruction relationship of pathogens for aerated piles (43).

Odor control is the primary environmental consideration in the operation of an aerated pile composting system. Good odor control results from prompt mixing of biosolids and bulking agent and formation of the aerated pile. In addition, lumps of material or puddles of liquid must not be allowed to remain in the mixing area. No thin spots or holes should be present in the compost blanket. There should be leak-proof transport of aeration air between blower

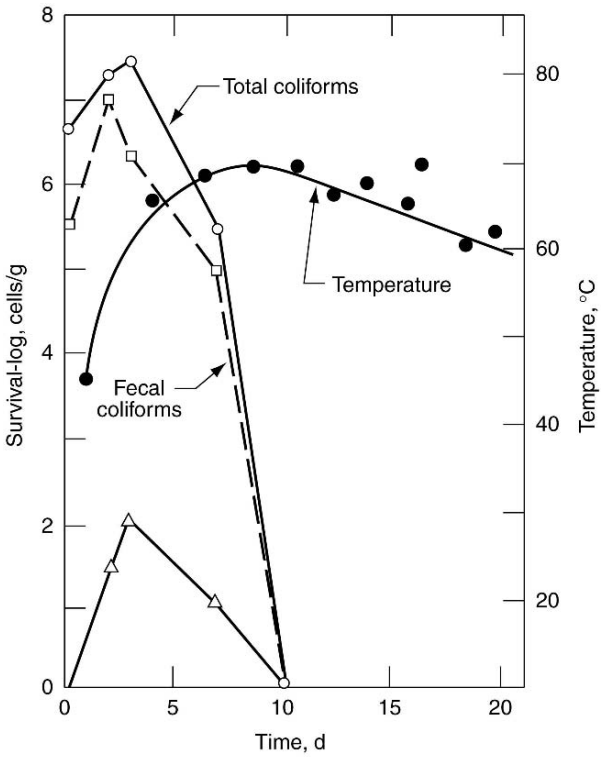


Fig. 16.9. Destruction of pathogenic organisms as a function of time and temperature during composting of undigested biosolids by the aerated pile method (15).

and odor filter pile. Moisture content within odor filter piles should be kept below 50 percent. Condensate, leachates, and runoff from the piles must be collected and treated as quickly as possible. The compost should be adequately cured before it is removed from the area, and any unstabilized material should be recycled back into the composting process for further treatment (56, 57).

8. IN-VESSEL COMPOSTING SYSTEM

8.1. Process Description

In-vessel composting occurs within a contained vessel, enabling the operator to maintain closer control over the process in comparison with other composting methods. The in-vessel systems are designed to minimize odors and process time by controlling environmental conditions such as air flow, temperature, and oxygen concentration. A typical flow diagram for in-vessel composting is shown in Figure 16.10. A mixture of dewatered wastewater solids and bulking agent is fed into a silo, tunnel, channel, or vessel. Augers, conveyors, rams, or other devices are used to aerate, mix, and move the product through the vessel to the discharge

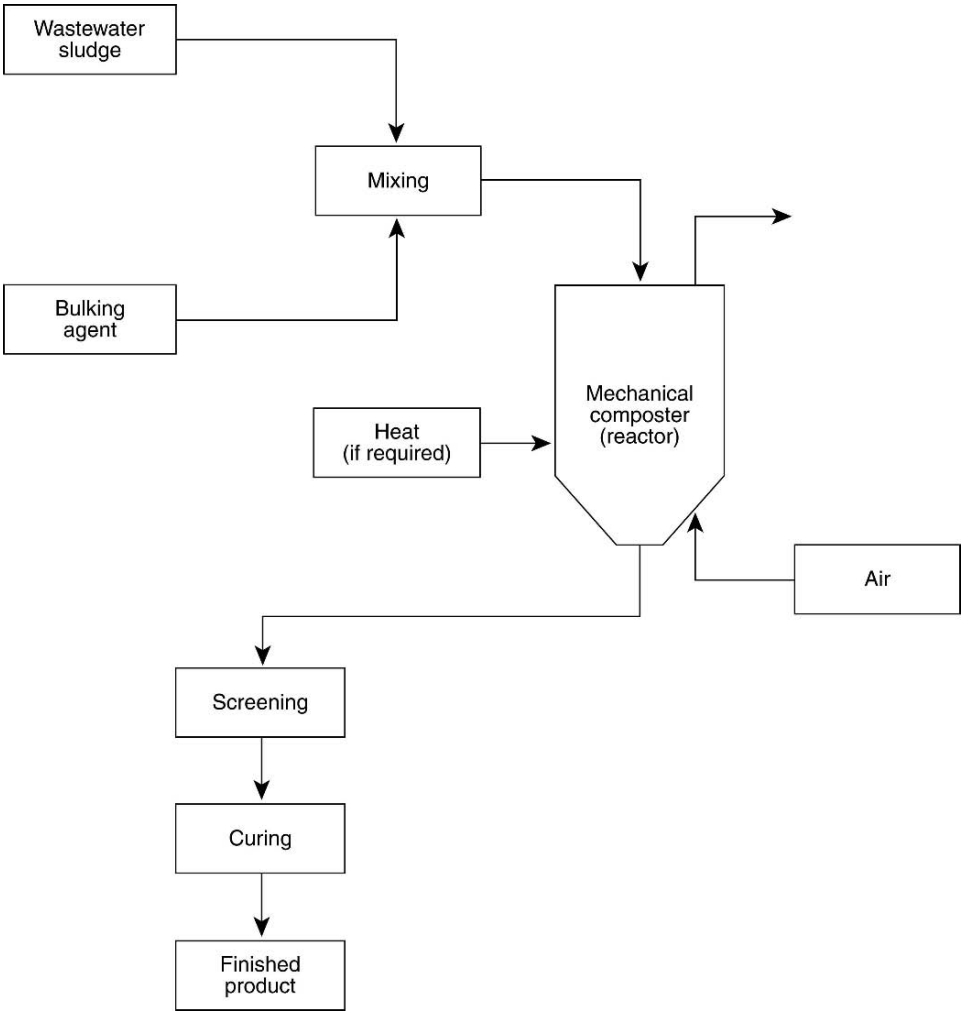


Fig. 16.10. Process flow diagram for confined composting system (15).

point (1). Air is generally blown into the mixture. After active composting, the finished product is usually stored in a pile for additional curing prior to distribution.

There are several types of in-vessel composting reactors: vertical plug-flow and horizontal plug-flow shown in Figure 16.11, and agitated bins shown in Figure 16.12. The primary difference involves the aeration systems and loading/unloading provisions. The first two systems operate as plug-flow, which means that biosolids and bulking agent are loaded on a periodic basis (typically daily or weekly) while the product compost is discharged from the opposite end of the system on roughly the same schedule (18). The vessel is only completely emptied for maintenance. A typical composting vessel is shown in Figure 16.13.

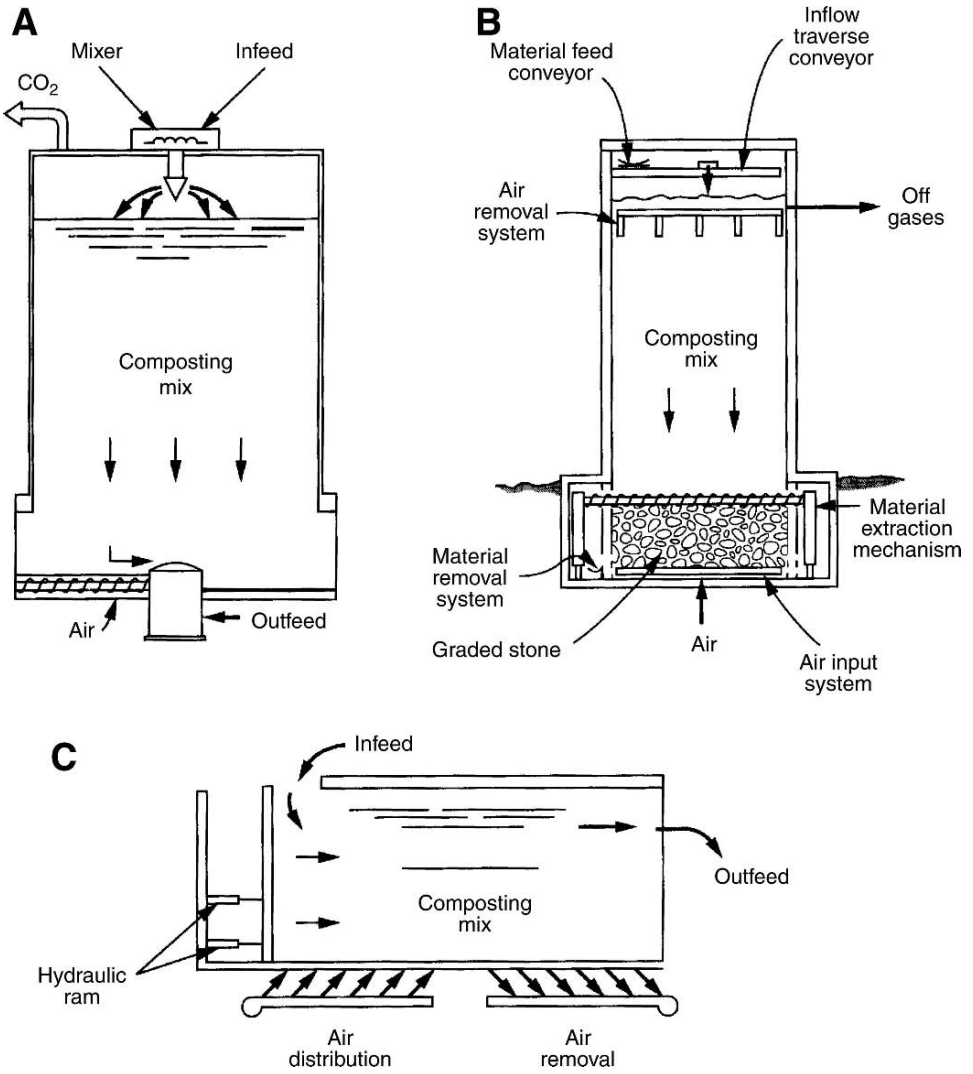


Fig. 16.11. Plug-flow in-vessel composting bioreactors (66): (A) cylindrical; (B) rectangular; (C) tunnel.

In vertical plug-flow systems, the biosolids and bulking agent mixture is introduced into the top of the reactor vessel and compost is discharged out the bottom by a horizontally rotating screw auger. Air is introduced in these systems either from the bottom and travels up through the composting mass where it is collected for treatment or through lances hanging from the top of the reactor.

In horizontal plug-flow systems, the compost and bulking agent mixture is loaded into one end of the reactor. A steel ram pushes the mixture through the reactor. Air is introduced and

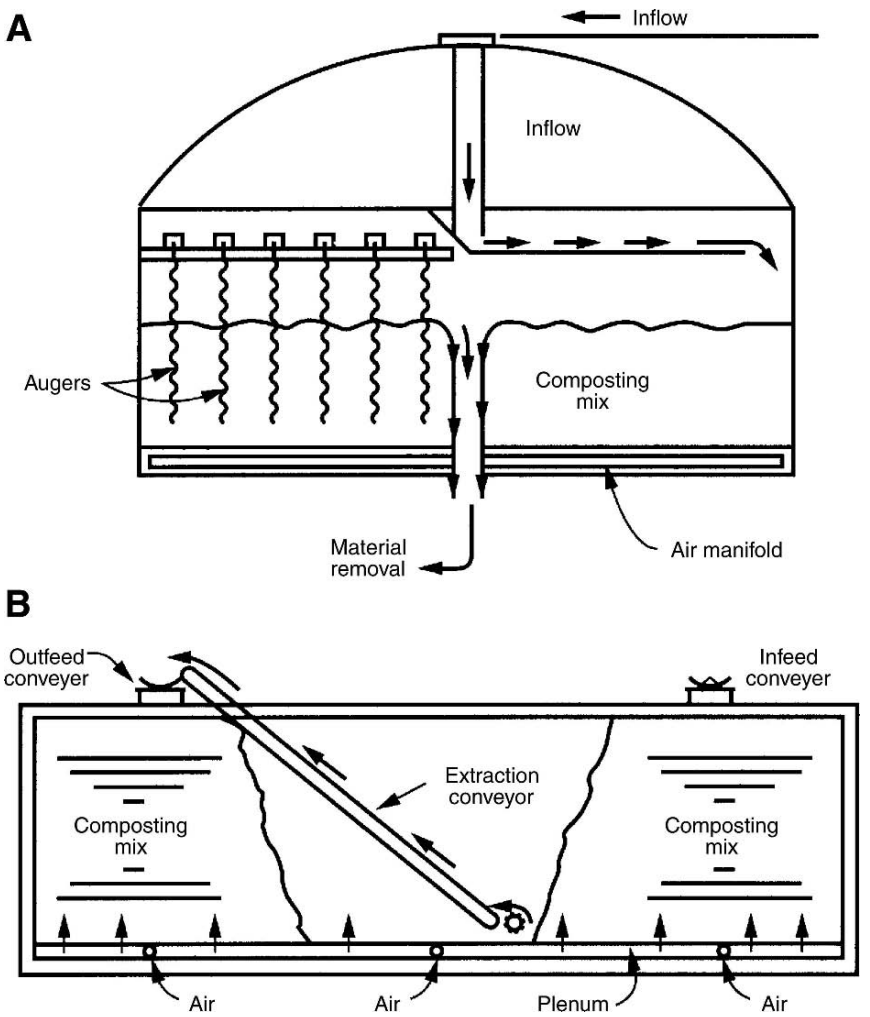


Fig. 16.12. Agitated (mixed) in-vessel composting bioreactors (66): (A) circular; (B) rectangular.

exhausted through slots in the floor of the reactor. Compost is discharged from the end of the reactor opposite the ram.

The agitated bed reactors are typically open topped. The biosolids and bulking agent mixture is loaded from above. The composting mass is periodically agitated using a mechanical device and air is introduced through the floor of the reactors. Agitated bed reactors can be operated as either plug flow or batch operations. In batch operations, the vessel is loaded with biosolids and bulking agent, processing takes place, and the vessel is emptied.

An odor control system is an inherent part of in-vessel design. The cost of an odor control system can account for up to 50 percent of both capital and operation and maintenance costs (18). Composting facilities usually use either wet scrubbers or biofilters for odor control.

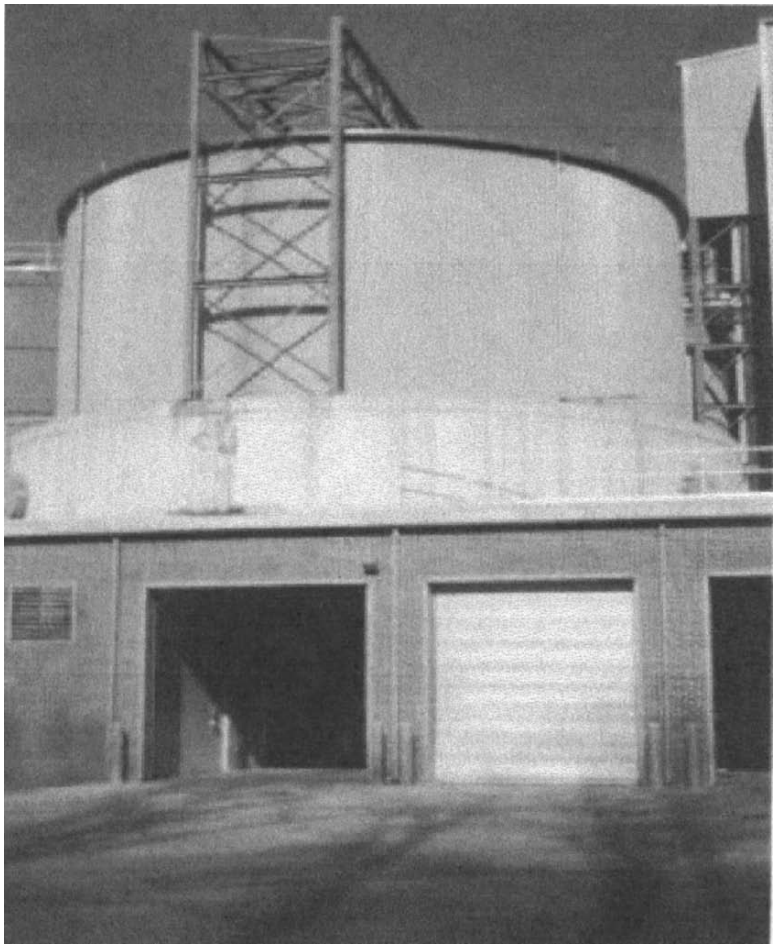


Fig. 16.13. Cylindrical composting vessel (1).

The level of odor control required is a function of the quality and quantity of air to be treated, the results of air dispersion modeling, and proximity to occupied dwellings (24).

In-vessel systems are designed for 10 to 21 days of active composting. Some state regulations dictate detention times for composting systems. The detailed design criteria for in-vessel systems can be found in *Composting Engineering* (58).

8.2. Advantages and Disadvantages

Shorter detention times, 14 days instead of the 20 days used in the unconfined systems, are usually specified by in-vessel equipment manufacturers (59). In-vessel technology offers the following advantages (11, 18):

1. The composting process can be more closely controlled
2. The effects of weather are diminished

3. Less bulking agent may be required
4. The quality of the resulting product is more consistent
5. Less manpower is required to operate the system and staff is less exposed to the composting material
6. Process air can be more easily collected for treatment to reduce odor emissions
7. Less land area is required
8. Public acceptance of the facility may be better

There are also disadvantages associated with in-vessel composting which must be considered before selecting this technology for wastewater solids management. In-vessel composting is generally more costly than other composting methods, particularly with respect to capital expenditures. In addition, because it is more mechanized, more equipment maintenance is necessary.

8.3. Applicability

In-vessel technology is more suitable than other composting technologies in suburban and urban settings because the system allows for containment and treatment of air to remove odors before release. The requirement for a relatively small amount of land also increases its applicability in these settings over other types of composting. The market for use of the resulting product will generally be more readily available in suburban and rural areas rather than urban settings (18). However, the usefulness of the final product in home gardening and commercial operations makes the material marketable in urban as well as rural areas. This is especially true for good quality material that does not emit foul odors (1).

Another important consideration before selecting the technology to be used for composting is the availability of adequate and suitable manpower. Composting is typically labor-intensive for the following reasons:

1. Bulking agents must be added.
2. Turning, monitoring, or process control is necessary.
3. Feed and finished materials must be moved with mechanical equipment.
4. Storage piles must be maintained for curing and distribution.
5. Bulking agents' recovery adds another step.

The number of operating in-vessel composting facilities for biosolids in the United States has steadily increased in the last two decades but has leveled off in recent years (18). According to a survey conducted in 1999, there were 54 in-vessel composting facilities processing wastewater residuals across the United States and 11 more facilities were in various stages of design or construction (7).

In-vessel mechanical processes are more capital-intensive than windrow and aerated static pile processes. This hinders the wide spread use of in-vessel composters and limit their application to cases in which the ultimate use of the compost is firmly established to justify the investment. Table 16.5 compares the three composting methods and highlights key features of each. Detailed information on composting can be found in Refs. (60–69).

Table 16.5
Comparison of composting methods (1)

Aerated Static Pile	Windrow	In-Vessel
Highly affected by weather (can be lessened by covering, but at increased cost)	Highly affected by weather (can be lessened by covering, but at increased cost)	Only slightly affected by weather
Extensive operating history both small and large scale	Proven technology on small scale	Relatively short operating history compared to other methods
Large volume of bulking agent required, leading to large volume of material to handle at each stage (including final distribution)	Large volume of bulking agent required, leading to large volume of material to handle at each stage (including final distribution)	High biosolids to bulking agent ratio so less volume of material to handle at each stage
Adaptable to changes in biosolids and bulking agent characteristics	Adaptable to changes in biosolids and bulking agent characteristics	Sensitive to changes in characteristics of biosolids and bulking agents
Wide-ranging capital cost	Low capital costs	High capital costs
Moderate labor requirements	Labor intensive	Not labor intensive
Large land area required	Large land area required	Small land area adequate
Large volumes of air to be treated for odor control	High potential for odor generation during turning; difficult to capture/contain air for treatment	Small volume of process air that is more easily captured for treatment
Moderately dependent on mechanical equipment	Minimally dependent on mechanical equipment	Highly dependent on mechanical equipment
Moderate energy requirement	Low energy requirements	Moderate energy requirement

9. COSTS

The capital costs of aerated static pile or windrow configuration may be lower than in-vessel composting configurations, but costs increase markedly when cover is required to control odors. More highly mechanized in-vessel systems are often more costly to construct, but tend to be less labor intensive. On the other hand, in-vessel systems tend to be less flexible in their ability to adapt to changing properties of biosolids and bulking agent feedstocks (1).

Capital costs of in-vessel systems range from USD 42,500 to 108,000/T (USD 39,300 to 97,000/t) per day processing capacity. A typical aerated static pile facility costs approximately USD 42,500/T (USD 39,300/t) per day of processing capacity (70, 71).

Typical operation and maintenance (O&M) costs for in-vessel systems range from USD 196/t per day to greater than USD 262/t per day. Aerated static pile O&M costs average USD 196/t per day (70, 71). Costs for windrow systems fall between the costs for in-vessel and aerated static pile. The selling price for compost ranges from USD 6.5 to 13/yd^3 or USD 13 to 26/t. Some facilities allow landscapers and homeowners to pick up compost for little or no charge.

Cost estimates were updated from 1994 to reflect the 2008 costs using the Cost Index for Utilities (Appendix A); all costs were multiplied by a factor of $552.16/424.91 = 1.30$ (72).

For further details, the reader is referred to the extensive literature dealing with this subject (11, 15, 59, 73–75).

10. DESIGN EXAMPLES

10.1. Design Example 1-Windrow Process

This design example illustrates the procedure for a 10 MGD (0.45 m^3/s) municipal wastewater secondary treatment plant. The dewatered, digested primary and secondary biosolids (20 percent solids) is generated at the rate of one dry ton/MG (0.00024 T/m^3). The compost facility will handle 10 dry ton/d (9 T/d) at 20 percent solids, seven days per week. The values for the process design variables are similar to those reported for Beltsville. The availability and cost of amendments and suitable land for the operation will strongly influence the economic analysis of the project. This design example, however, does not consider these site-specific economic parameters (15).

The design of this windrow composting facility is based on the following assumptions:

(a) The water content and total weight of the compost mixture will be reduced by approximately 40 to 50 percent and volatile solids content will be reduced by about 20 to 40 percent. The density will decrease by 15 to 25 percent because of evaporation.

(b) The values for the process variables are assumed to be as follows:

$$\begin{array}{llll} S_C = 0.20 & S_R = 0.70 & S_A = 0.90 & S_M = 0.40 \\ V_C = 0.50 & V_R = 0.35 & V_A = 0.90 & V_M = 0.50 \\ k_C = 0.45 & k_R = 0.15 & k_A = 0.50 & \end{array}$$

(c) If the mixture has a high ratio of water to degradable organics by weight (W ratio greater than ten), amendment will be added to reduce W.

Solution

The amount of finished compost to be recycled can be calculated using Equation (3).

$$\begin{aligned} X_R &= X_C(S_M - S_C)/(S_R - S_M) \\ X_R &= 50(0.04 - 0.20)/(0.70 - 0.40) \\ X_R &= 33.3\,\text{t/d}(30.3\,\text{T/d}) \end{aligned} \tag{3}$$

This indicates that if a mixture moisture content of 40 percent is to be obtained, 0.67 t of finished compost must be added to each ton of biosolids cake to be composted.

The ratio W is checked using Equation (4) in order to determine whether to compost.

$$W = [X_C(l - S_C) + X_R(l - S_R)]/[X_C S_C V_C k_C + X_R S_R V_R k_R] \quad (4)$$

$$W = [50(1 - 0.20) + 33.3(1 - 0.70)]/[50(0.20)(0.50)(0.45) + 33.3(0.70)(0.35)(0.15)]$$

$$W = 14.4$$

The calculated value for W is too high, indicating that amendment addition is required. Increasing the recycle rate to create a mixture of 50 percent solids ($X_R = 50$ t/d) would only lower W to 13.5, because the proportion of degradable organics does not increase significantly in the mixture.

Assuming that 1.0 t (0.9 T) amendment per ten t (9 T) of biosolids cake is added to the mixture, the recycle rate can be calculated using Equation (5):

$$X_R = [X_C(S_M - S_C) + X_A(S_M - S_A)]/(S_R - S_M) \quad (5)$$

$$X_R = [50(0.40 - 0.20) + 5(0.40 - 0.90)]/(0.70 - 0.40)$$

$$X_R = 25.0\,\text{t/d}(22.7\,\text{T/d})$$

The amount of recycled compost has dropped from 0.67 t/t (0.67 T/T) to 0.5 t/t (0.5 T/T) of biosolids cake. The ratio W is calculated using Equation (6):

$$W = \frac{[X_C(1 - S_C) + X_R(1 - S_R) + X_A(1 - S_A)]}{[X_C S_C V_C k_C + X_R S_R V_R k_R + X_A S_A V_A k_A]} \quad (6)$$

$$W = \frac{[50(1 - 0.20) + 25(1 - 0.70) + 5(1 - 0.90)]}{[50(0.20)(0.50)(0.45) + 25(0.70)(0.35)(0.15) + 5(0.90)(0.90)(0.50)]}$$

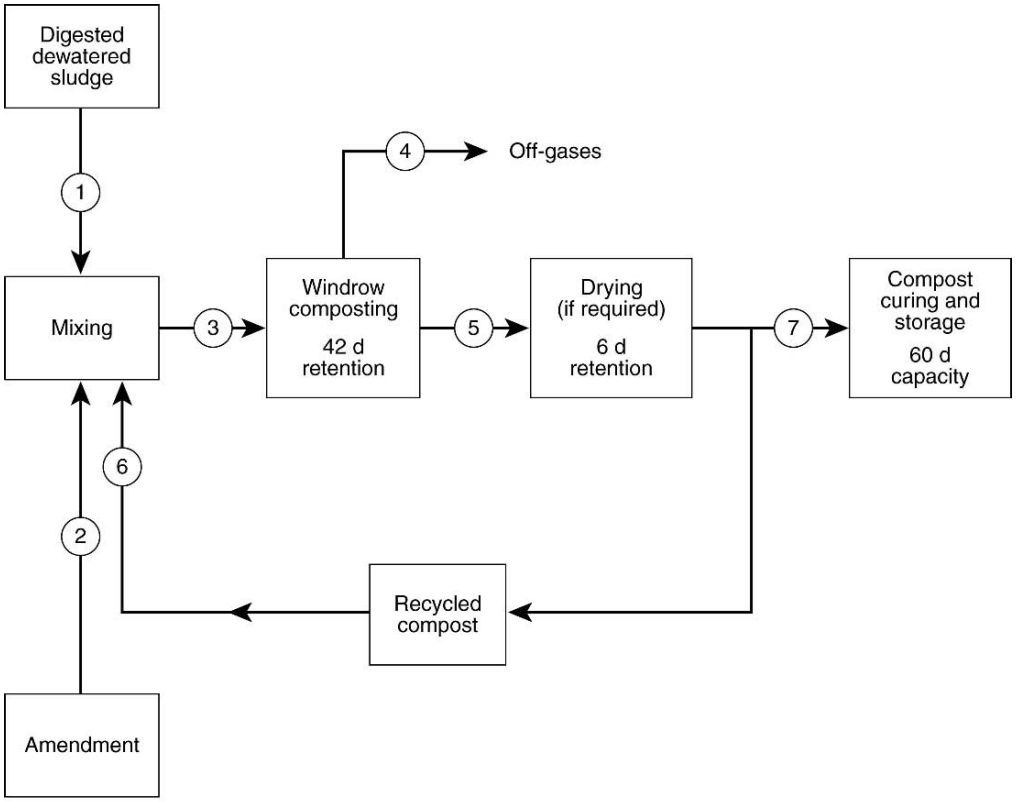
$$W = 9.2$$

This mixture of biosolids cake, recycled compost, and amendment is self-sustaining and will degrade properly. Figure 16.14 illustrates this process and shows the materials balance.

A 7-ft (2 m) high, 65-ft (20 m) long, windrow with a base of 15 ft (4.6 m) is constructed each day. Longer windrows can be made if the windrow is extended each day with the mixture to be composted. The final volume of composting at the end of six weeks of turning is approximately 65 percent of the original volume. In continuous operation there would be about 11 windrows, 250-ft (76 m) long.

Each windrow must be turned at least two times per day for the first five days to mix the materials completely, to minimize odors, and to insure sufficient oxygen transfer. After the initial five-day period, the windrows must be turned frequently enough to maintain the proper oxygen level and temperature in the composting material. This is dependent on weather conditions.

Other site operations must include a mixing area, maintenance and operations building, a curing area to stockpile the finished compost, and enough land area for handling all other site operations and for future expansion.



7 (d/wk) operation

Location	Wet (t)	Percent solids	Dry (t)	Density (lb/yd³)	Volume (yd³)	Percent volatile solids
1	50	20	10.0	1600	63	50
2	5	90	4.5	1000	10	90
3	80	40	32.0	1300	123	50
4	41	–	5.0	–	–	–
5	39	70	27.0	1000	78	35
6	25	70	17.5	1000	50	35
7	14	70	9.5	1000	28	35

Fig. 16.14. Process flow diagram-windrow composting of biosolids from a 10 MGD activated sludge plant (15).

Equipment required for the operation includes a windrow turning machine; a front-end loader for site preparation, dismantling piles and loading transfer trucks; and transfer trucks to haul the biosolids and amendment to the compost facility and to haul the finished compost away.

Optimum windrow compost design will do the following:

1. Minimize hauling and handling cost
2. Maximize use of existing equipment in the compost operation

Table 16.6
Densities of compost and bulking agents used in Example 2 (15)

	Bulk Density	
Constituent	lb/yd^3	kg/m^3
Dewatered Sludge (20% solids)	1,600	960
New Wood Chips	500	300
Recycled Wood Chips	600	360
Screened Compost	865	519
Unscreened Compost	1,000	600

3. Minimize the use of amendment which adds to the cost and is not recoverable
4. Maximize the solids content of the dewatered digested sludge cake to minimize the amount of recycled compost used for moisture control and also reduce the amount of amendment required. The cost of dewatering should not exceed the savings at the compost facility.

10.2. Design Example 2-Extended Aerated Pile System

This design example is based on a Beltsville-type biosolids composting system utilizing existing technology and available design criteria. The example provided is specific to a 10 MGD (0.45 m^3/s) municipal wastewater secondary treatment plant (15).

The weight and volume of biosolids and bulking agent at various points in the process must be known so that the volumetric flow capacity of a composting facility can be determined. The basic design decisions to be determined are:

1. The bulking agent to biosolids ratio, and
2. The ratio of new to recycled bulking agent

The materials balance in this example is based on the following assumptions:

1. Biosolids to be composted are 50 wet t/d (45 T/d) of undigested biosolids, seven days per week, with no digestion
2. Wood chips are added to the wet biosolids at the rate of 2 yd^3/yd^3 (2 m^3/m^3) of wet biosolids
3. Three-fourths of the chips are recovered by screening and reuse
4. The water content and total weight of the compost mixture is reduced by approximately 30 to 40 percent and volatile solid's content is reduced by about 10 to 15 percent. The density decreases 15 to 20 percent because of evaporation
5. The extended aerated pile system will be used

Information on the bulk density of biosolids is surprisingly scarce. Tests conducted at Be1tsville for an engineering study of a large-scale composting facility provide some basic data on the bulk density of biosolids and wood chip bulking agents. The bulk densities used in this design example are shown in Table 16.6:

It is also assumed that the process variables have the following values:

$$\begin{array}{lll} S_C = 0.20 & S_B = 0.70 & S_R = 0.70 \\ V_C = 0.75 & V_B = 0.90 & V_R = 0.80 \\ k_C = 0.45 & k_B = 0.10 & k_R = 0.10 \end{array}$$

Solution

(a) The bulking agent to biosolids ratio
Biosolids composting will operate 5 days per week, 8 hours per day using the extended aerated static pile method. The volume to be composted per work day is as follows:

$$(50\,\text{wet t/week-d})\,(7\,\text{week-d/5 work-d}) = 70\,\text{wet t/work-d}\,(63.5\,\text{T/work-d})$$

It is assumed that the dewatered biosolids arrive onsite 5 d/week from the dewatering operation which runs only 5 d/week. Equalization storage to cover weekend operation of the plant is provided for biosolids in the liquid state upstream from the dewatering process.

The amount of recycled and new wood chips can be calculated using Equations (7) and (8) and assuming:

$$f_1 = 0.75, \text{ and}$$

$$f_2 = 0.25.$$

$$X_R = (1 - f_2)\, f_l X_C \tag{7}$$

$$X_B = f_1 X_C - X_R \tag{8}$$

$$X_R = (1 - 0.25)(0.75)(70) = 39.4\,\text{t/d}(35.7\,\text{T/d})$$

$$X_B = (0.75)(70) - 39.4 = 13.1\,\text{t/d}(11.9\,\text{T/d})$$

The ratio W can be calculated using Equation (2):

$$W = \frac{[X_C(l - S_C) + X_A(l - S_A) + X_B(l - S_B) + X_R(l - S_R)]}{[X_C S_C V_C k_C + X_A S_A V_A k_A + X_B S_B V_B k_B + X_R S_R V_R k_R]} \tag{2}$$

$$W = \frac{[70(1 - 0.2) + 39.4(1 - 0.7) + 13.1(1 - 0.7)]}{[70(0.2)(0.75)(0.45) + 39.4(0.7)(0.9)(0.1) + 13.1(0.7)(0.8)(0.1)]}$$

$$W = 9.0$$

Since W is less than 10, no amendment addition is required.

(b) The ratio of new to recycled bulking agent
The daily volume of the compost material is calculated using the assumed values previously stated:

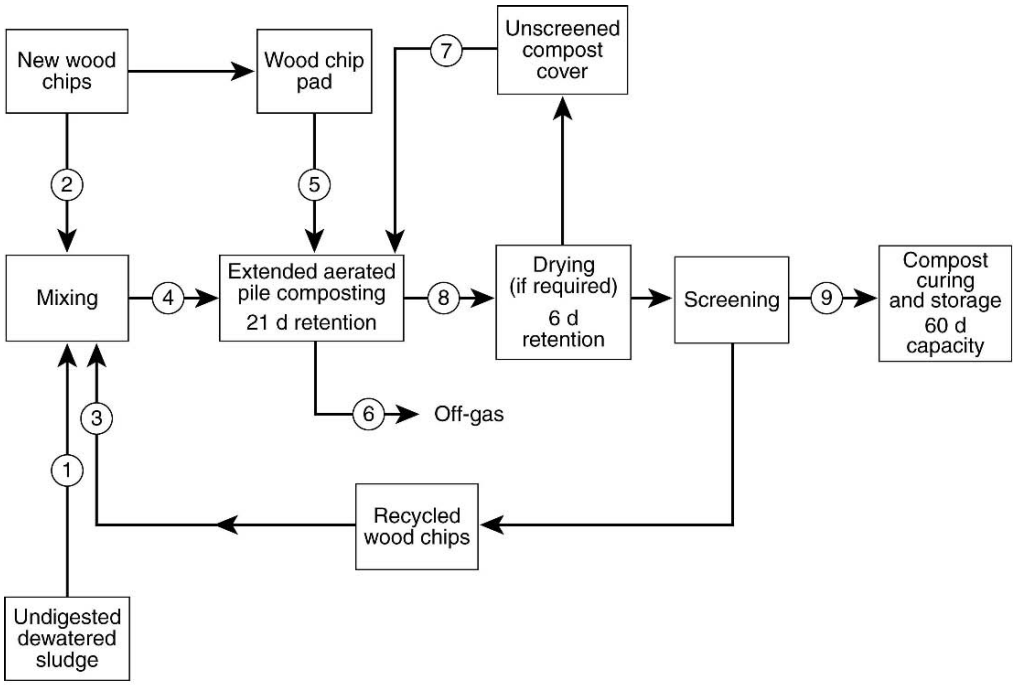
Constituent	Mass (t/d)	Volume (yd^3/d)
Dewatered biosolids	70	87.5
New wood chips	13.1	52.4
Recycled wood chips	39.4	131.3
Total	122.5 (111.1 T/d)	271.2 ($206.8\,m^3$/d)

The pile will be 8 ft (2.4 m) high and 50 ft (15 m) long. Each day, the pile will be extended 18.5 ft (5.6 m). The amount of new wood chips required to construct a one-foot (0.3 m) thick pad for the compost is as follows:

$$(50\,\text{ft})(18.5)(1\,\text{ft})/27\text{ft}^3/\text{yd}^3 = 34.3\,\text{yd}^3/\text{d}(26.2\,\text{m}^3/\text{d})$$

Unscreened compost is required each day to cover the pile. This layer will be 18 inches (0.46 m) thick:

$$(50\,\text{ft})(18.5\,\text{ft})(1.5\,\text{ft})/27\,\text{ft}^3/\text{yd}^3 = 51.4\,\text{yd}^3/\text{d}(39\,\text{m}^3/\text{d})$$



5 (d/wk) operation

Location	Wet (t)	Percent solids	Dry (t)	Density (lb/yd3)	Volume (yd3)	Percent volatile solids
1	70	20	14	1600	87.5	75
2	13.1	70	9.2	500	52.4	90
3	39.4	70	27.6	600	131.3	80
4	122.5	41	50.8	900	271.2	80
5	8.6	70	8	500	34.3	90
6	58.7	–	10.3	–	–	–
7	18.8	65	12	725	51.4	65
8	90	65	58.5	725	248.3	65
9	32	60	18.9	975	54.6	45

Fig. 16.15. Process flow diagram for the extended pile compost facility for a 10 MGD activated sludge plant (15).

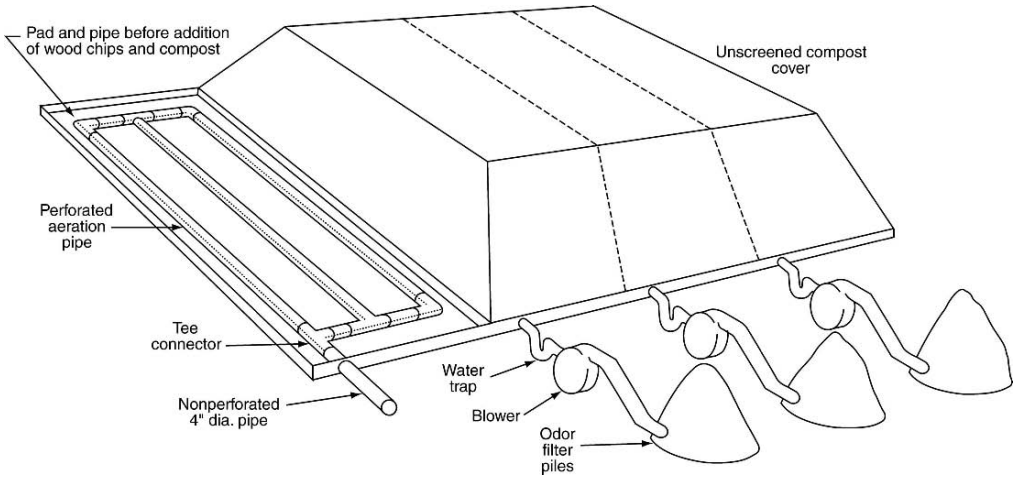


Fig. 16.16. Design example-extended aerated pile construction (15).

Figure 16.15 is the process flow diagram for the extended aerated pile compost facility and summarizes the design materials balance.

Approximately 250 ft (76 m) of 4-inch (10-cm) diameter perforated aeration pipe, 50 ft (15 m) of non-perforated pipe, three 4-inch (10-cm) tee connectors, and one blower/timer unit with weather protection and condensate collection system are required for each daily pile. Only one blower rated at 335 ft^3/min (158 L/s) will be used to draw air into the pile. In general, the blower should be rated at a minimum of 150 ft^3/h/wet ton (1.3 L/s/T) of biosolids in the daily pile. Non-perforated pipe should be used to connect the aeration pipe loop to the blower. The exhausted air will be filtered in a pile of screened compost. The filter pile will contain at least one yd^3 of material/30 wet ton (1 m^3/35.5 T) of biosolids in the daily pile or 4 yd^3 (3 m^3) for this design. Figure 16.16 illustrates this design example. The minimal area requirements for various composting site components are shown in Table 16.7.

The overall space required is about 3 acres (1.2 ha) which is 0.15 acres/t/d (0.07 ha/T/d) of dry biosolids composted. Reducing the bulking agent would decrease the area required.

Although porosity is the key factor for the aerated pile, control of moisture is important for a successful biosolids composting system. The biosolids should be dewatered or mixed with sufficient bulking agent to obtain enough porosity in the composting piles for optimum composting conditions. For optimum composting the composted mixture should have a solids content of not less than 40 percent or more than 50 percent.

Approximately 8.5 ft^3 of air/min/ton of dry solids (4 L/s/T) in the pile is required. At Beltsville, this was delivered by a centrifugal fan operating at 5 inches differential water pressure (1.25 kN/m^2) (41). The Bangor, Maine system uses a 1/3 hp (0.25 kw) blower rated at 335 ft^3/min (158 L/s) at 5 inches water pressure (1.25 kN/m^2) for each pile consisting of 50 yd^3 (38 m^3) biosolids and 150 yd^3 (114 m^3) bulking agent (8).

The blowers are operated intermittently to maintain the oxygen level in the 5 to 15 percent range and to obtain as uniform a temperature as possible.

For large composting systems, a permanent central blower system may be considered. A header pipe could be utilized to provide the necessary suction for each pile. Only one or two large blowers located in a covered area would be required. Although capital cost would be high because of the needed piping and control devices, the operation and maintenance costs of

Table 16.7
Minimal composting area requirements for Example 2 (15)

Minimal Composting Area Requirements 50 wet tons per day (45 T/day) 10 dry tons per day (9 T/day)		
	Area Required	
Function	ft^2	m^2
Truck unloading and mixing	5,000	465
Composting		
(28 days) (50) (18.5) (1.15 excess)	30,000	2,792
Unscreened compost	10,000	931
Drying and screening	20,000	1,862
Compost curing and storage		
(60 day) (200 yd^3/day) (27 wet tons)		
(10 ft deep) + excess	33,000	3,071
New wood chip storage		
(60 day) (87 yd^3/day) (27 wet tons)		
(12 ft deep) + excess	15,000	1,396
Subtotal	113,000	10,517
Maintenance building, operations building and laboratory, Lunch room and locker room	4,000	372
Employee and visitor parking	5,000	465
Miscellaneous storage	1,000	93
Subtotal	10,000	930
Total	123,000	11,447

Land Utilization = 6.6 dry tons per acre (1.48 T/ha).

many individual blowers would be eliminated. On the other hand, a central blower system is not especially flexible. Since it is important to maintain the proper aeration rates in each pile, an air flow metering device will be required for each pile. A decision for or against a permanent system would be based on economic analysis and the need for system flexibility to handle changing composting conditions.

The composting area should be paved. Probably the most efficient design in a permanent facility involves the use of fixed aeration and drainage systems. The aeration piping and drainage system could be placed in trenches in the composting pad and the blowers placed in permanent protected structures and equipped with water traps and controls. The disadvantages of this type of combined system are the high initial cost and the reduced flexibility of operation. Possible elimination of the one-ft (0.3 m) wood chip pad and the disposable plastic pipe processed through the screens is a potential advantage of fixed trenches for the aeration pipes. Special precautions would be necessary to keep the centralized aeration piping and pile drainage trenches from clogging and to provide for condensate water drainage.

Odor filter piles should be replaced periodically. The filter piles are replaced every other month at Bangor; during cool weather the system has operated without significant odor problems and

with no filter piles. At Beltsville, the odor filter pile is replaced each time the compost pile is dismantled.

After the piles are formed, they should be covered with a layer of compost or wood chips for insulation and to prevent the dust which is caused by excessive drying of the outer pile edges from blowing.

Most composting facilities use a base layer of bulking agent or unscreened compost to cover the aeration piping. However, the piles are constructed at Bangor with no special base layer; the biosolids-bulking agent mixture is placed directly on the aeration piping.

NOMENCLATURE

X_C = Total wet weight of sludge cake produced/d.
X_A = Total wet weight of amendment/d.
X_R = Total wet weight of recycle/d.
X_B = Total wet weight of external bulking agent/d.
X_M = Total wet weight of mixture/ d.
S_C = Fractional solids content of sludge cake (0.20 to 0.55).
S_A = Fractional solids content of amendment (0.50 to 0.95).
S_R = Fractional solids content of recycle (0.60 to 0.75).
S_B = Fractional solids content of external bulking agent (0.50 to 0.85).
S_M = Fractional solids content of mixture (0.40 to 0.50).
V_C = Volatile solids content of sludge cake, fraction of dry solids (0.40 to 0.60) for Digested; (0.60 to 0.80) for Raw.
V_A = Volatile solids content of amendment, fraction of dry solids (0.80 to 0.95).
V_R = Volatile solids content of recycle, fraction of dry solids (0.00 to 0.90).
V_B = Volatile solids content of external bulking agent, fraction of dry solids (0.55 to 0.90).
V_M = Volatile solids content of mixture, fraction of dry solids (0.40 to 0.80).
k_C = Fraction of sludge cake volatile solids degradable under composting conditions (0.33 to 0.56).
k_A = Fraction of amendment volatile solids degradable under composting conditions (0.40 to 0.60).
k_R = Fraction of recycle volatile solids degradable under composting conditions (0.00 to 0.20).
k_B = Fraction of external bulking agent volatile solids degradable under composting conditions (0.00 to 0.40).
k_M = Fraction of mixture volatile solids degradable under composting conditions (0.20 to 0.60).
t = ton (english)
T = Ton (metric)

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APPENDIX

United States Yearly Average Cost Index for Utilities U.S. Army Corps of Engineers (72)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

17
Vermicomposting Process

Lawrence K. Wang, Yung-Tse Hung, and Kathleen Hung Li

CONTENTS

Abstract Vermistabilization, also known as vermicomposting, is a biodegradation process for stabilization of biosolids and organic solid wastes using earthworms. The worms maintain aerobic conditions in the organic substances and accelerate and enhance the biological decomposition of the organic substances. This chapter introduces and reviews the vermistabilization process. The technology development, technical problems, legal problems, technology breakthrough, current status, available resources, engineering design, and recent advances of the process are discussed in detail.

Key Words vermistabilization • vermicomposting • earthworms • composting • biosolids • organic solid wastes • sludges • sludge disposal • biosolids management • solid waste disposal • engineering design • flow diagram.

1. INTRODUCTION

1.1. Summary

Vermicomposting is a novel municipal biosolids and solid waste treatment process that uses earthworms (Oligochaete annelids) for the biodegradation of the biosolids and/or solid waste. This system is alternately called earthworm conversion, Vermicomposting, vermistabilization, worm composting or annelidic consumption. The worms maintain aerobic conditions in

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the organic substances while accelerating and enhancing the biological decomposition of the organic substances. The main product of the Vermicomposting (earthworm conversion) process is the worm's castings. In some process arrangements, there may be a net earthworm production. The excess earthworms may then be sold for fish bait or animal protein supplement. Earthworm marketing is a complex problem; for municipal biosolids applications, surplus earthworms may be considered a by-product while the principal product is the castings, which can be a resource.

This chapter presents the following: (a) an introduction and review of the Vermicomposting process; (b) technology development, technical problems, legal problems, and technology breakthrough of the process; (c) current status and resources; (d) Vermicomposting process design considerations; (e) process applications; and (f) future development and directions of the process.

1.2. Process Description

Vermicomposting differs from the conventional composting of wastewater treatment plant biosolids. In the Vermicomposting process, worms are used to develop an optimum environment for consuming or metabolizing the biosolids and producing feces or castings. These castings may be used as a soil conditioner (1–38, 39–63). In the conventional composting process, microorganisms are used for the degradation of biosolids and other putrescible organic solid materials under an aerobic metabolism environment. Conventional composting is also suitable for converting undigested primary/secondary biosolids, and certain solid wastes into an end product amenable to resource recovery with a minimum capital investment and relatively small operating commitment (39, 40, 43).

Figure 17.1 shows a basic simple Vermicomposting process (59, 62) that requires worm beds and an ample supply of worms. Generally, digested and dewatered biosolids are put into the beds, although experiments are underway, where raw liquid sludge is placed in beds. If anaerobic digestion is used prior to earthworm conversion, additional pretreatment may be needed. A bulking agent such as wood chips may be useful in some cases for keeping the bed porous and aerobic, especially if moisture is high. Biosolids are, however, generally applied without any bulking agent. A worm bed may take the form of a simple tray. Windrows similar to those for composting may also be used. After the worms have consumed the biosolids, they must be separated from the castings. This may be done with an earthworm harvester, a drum screen that rotates on a nearly horizontal axis. Castings fall through the screen openings while worms tumble through the length of the drum. Section 5 contains some critical operational parameters for the earthworm conversion process.

2. TECHNOLOGY DEVELOPMENT

Conversion of sludges (or biosolids) into topsoil by earthworms was initially attempted by Mitchell et al. of the State University of New York at Syracuse, College of Environmental Science and Forestry in 1977. (1). Later, Mitchell et al. investigated the potential role of the earthworm, *Eisenia foetida*, on the decomposition of sewage biosolids in drying beds and reported the results in 1980 (2). Specifically, Mitchell et al. sought to determine the

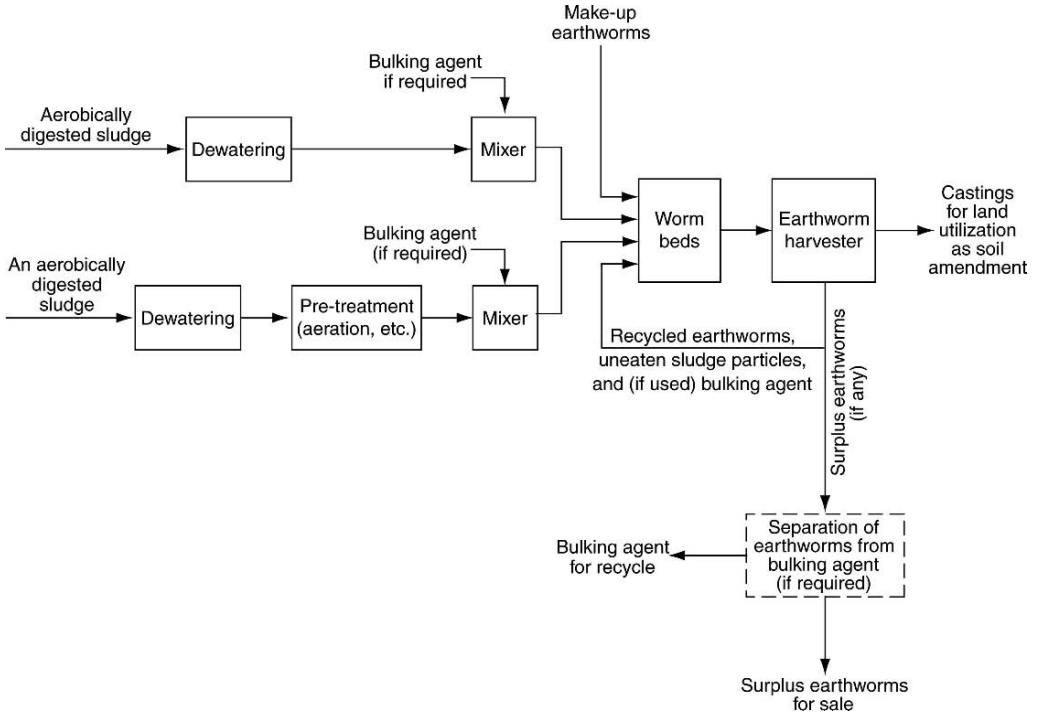

Fig. 17.1. Diagram of an earthworm process (*Source:* US EPA).

decomposition rates of biosolids in drying beds as indexed by consumption of oxygen and evolution of carbon dioxide and methane, to ascertain whether *E. foetida* can alter the form and rate of decomposition, and to ascertain the relationship among specific biotic and abiotic components in decomposition. At two facilities tested, the aerobic and anaerobic bacteria were abundant, and the dominant bacteria were not enteric. A computer simulation model regarding the role of macroinvertebrates in decomposition was used to analyze the effects of the earthworm.

In August 1980, Camp, Dresser and McKee, Inc., of Boston, MA, U.S.A, completed a technical report (3) which assessed the technical and economic feasibility of Vermicomposting or vermistabilization process based on several pilot-scale studies conducted by private entrepreneurs. The assessment was based on examining facilities and costs for a municipal operation serving (a) a community of 50,000 persons and (b) a community of about 500,000 persons. Vermicomposting was compared to three other methods of solid waste management: sanitary landfill, windrow composting, and combustion. In 1980, Vermicomposting was estimated to cost about \$24 to \$32 per ton of waste processed. (Note: 1 ton = 2,000 pounds; 1 pound = 0.454 kilograms).

In 1981, Hornor and Mitchell (4) studied the effect of the earthworm, *Eisenia foetida*, on fluxes of volatile carbon and sulfur compounds from sewage biosolids. Hartenstein (5) suggested the potential use of earthworms as a solution to biosolids management. In Hartenstein's

study at the State University of New York at Syracuse (5), the feasibility of using earthworms in management of municipal biosolidss was examined in detail. Results of tests performed by Hartenstein on two earthworm species – *E. eugeniae* and *E. foetida* – were reported. The following observations were made: (a) the toxicity of worm-casts to the earthworms signifies the need to retain *E. foetida* in its source of food (biosolids) as long as, or slightly longer than, the time required to convert the biosolids into castings; (b) knowledge of the quantity of material passing through the earthworm gut per unit of time, for a particular ingestible sludge, permits prediction of biosolids quantity manageable per unit time; and (c) *E. foetida* fails to gain weight rapidly, if at all, on unlimited supplies of certain organic materials.

Also in 1981, Collier and Livingstone (7) completed research sponsored by the National Science Foundation. They used earthworms of the redworm (*E. foetida*) species to accomplish vermicomposting, or vermistabilization of biosolids from the San Jose and Santa Clara Wastewater Treatment Plants in California, U.S.A Ninety tons of earthworm manure were produced from the biosolids over a five-year period. Different size windrows were populated with different densities of earthworms, and castings were harvested by passing windrow contents through a rotating screen which separated the worms from the castings for reuse. Plants in castings outgrew plants in topsoil by a factor of 4 to 1. Their 1981 cost analysis showed the system to be cost effective at a cost of $29.45 per dry ton in a 10 ton per day facility, and to return a profit of $3.34 per dry ton if castings were produced at the rate of 50 tons per day.

In 1982, Hartenstein (8) reported: (a) the metabolic parameters of the earthworm *Eisenia foetida* in relation to temperature, and (b) the potential use for manure management and as a source of protein biomass. In 1983, Chosson and Dupuy (9) demonstrated their improvement of the cellulolytic activity of a natural population of aerobic bacteria – enrichment culture, and presented their isolation and characterization of worm gut and compost cellulolytic strains. In 1984, Hartenstein et al. (10) attempted to use earthworms in trickling filters for wastewater treatment.

In March 1984, Loehr et al. (11) presented the results of an investigation of the vermistabilization process using stabilized and unstabilized wastewater treatment sludges. Four earthworm species were evaluated: *E. foetida*, *E. eugeniae*, *P. hawayana*, and *P. excavatus*. *E. foetida* was found to have the greatest overall reproductive capacity. The best growth of *E. foetida* in terms of total biomass weight gain occurred in media that had a total solids content, wet basis, of between 9 and 17 percent. The best growth and cocoon production for this earthworm species was shown to occur at temperatures of 20°C to 25°C. With both dewatered and liquid biosolids, vermistabilization units functioned successfully for long periods of time – up to one year for dewatered biosolids and at least six months for the liquid sludges. Cost estimates indicated that the capital and annual costs of liquid vermistabilization were competitive with those for other sludge management systems.

In 1985, Loehr et al. of Cornell University (12) evaluated several fundamental factors that affect the performance of the vermistabilization process such as temperature, moisture content of the waste material and the combined use of several earthworm species (polyculture). The earthworms *Dendrobaena veneta, Eisenia foetida, Eudrilus eugeniae, Perionyx excavatus* and *Pheretima hawayana* were used in one or more of the studies. The best growth and reproduction of these species occurred at temperatures of 20 to 25°C. Growth of all five

species was reduced at 30°C and death occurred at 35°C. Of the five species, *Eisenia foetida* produced the largest number of young in a 20-week study. The growth of *Eisenia foetida* occurred optimally in media with a total solids content, wet basis, of between 9 to 16 percent. Polyculture did not exhibit any obvious advantages over monoculture.

Stabilization of liquid sludge, or biosolids, by vermistabilization process was also reported by Loehr et al. of the University of Texas at Austin, TX, U.S.A. (13). The investigators conducted basic studies to identify fundamental factors that affect the performance of the vermistabilization process and applied studies to determine design and management relationships. As earthworms are a key component of the liquid sludge vermistabilization (LSVS) process, control reactors that did not contain worms failed in a much shorter period of time than did the reactors with the worms. LSVS reactors that were not overloaded functioned successfully for 140 to 198 days and were stopped only because the project ended. Oxidized nitrogen (nitrates) in the drainage from the LSVS reactors indicated that aerobic conditions were being maintained. Liquid primary sludge and liquid wastes activated sludge (biosolids) can be stabilized by the LSVS process.

LSVS reactors were not adversely affected by short-term, large variations in loading rates. Liquid primary sludge was stabilized to about the same degree as liquid aerobically digested biosolids in the LSVS process. Moisture balances indicated an overall moisture loss of 4 to 20 percent. Loading rates of about 21,000 grams/sq. m/week volatile solids or less resulted in satisfactory operation of LSVS reactors stabilizing liquid primary and liquid waste activated sludge. Loading rates greater than 1200 grams/sq. m/week volatile solids could be used for LSVS reactors stabilizing liquid aerobically digested biosolids. With LSVS reactors, the disposal of residual stabilized solids occurs at long intervals. The total solids content of the stabilized residual solids in the LSVS reactors was from 14 to 24 percent, a considerable increase from the 0.6 to 1.3 percent that was added. LSVS proved to be a successful process for both dewatering and stabilization. The stabilized residual solids had approximately the same characteristics regardless of the type of liquid sludge added to the reactors. Size and cost estimates indicated that LSVS might be an economically feasible sludge management process.

Reviews of the literature on biosolids characteristics, solids concentration and conditioning, stabilization and inactivation, incineration, and ultimate disposal and utilization were conducted by Hasit of Weston, Inc., West Chester, PA, U.S.A. in 1985 (14) and 1986 (15). Vermistabilization was one of the biosolids management technologies reviewed and assessed.

In 1986, Stafford and Edwards (16) of Rothamsted Experimental Station, Harpenden, England used earthworms in the field to indicate levels of soil pollution and in the laboratory for the ecotoxicological testing of industrial chemicals. An earthworm bioassay procedure developed at the Waterways Experiment Station in Vicksburg, Mississippi, U.S.A., was modified and evaluated as a method of providing information on heavy metal bioavailability in contaminated soils and sediments from Europe. Eight soils/sediments containing elevated levels of at least one of the elements Zn, Cu, Cd and Pb were selected, as well as a control and a reference soil. Six earthworm species, including the WES bioassay earthworm *E. foetida*, and five field species, were grown in the soil for periods of 15, 28 or 56 days. Concentrations of the elements Zn, Cu, Cd, Ni, Cr, and Pb present in the earthworm samples (corrected for the

presence of soil-derived metals within the earthworm gut) were compared between earthworm species from the same soil and for each earthworm species from a range of metal contaminated soils/sediments.

A U.S. Patent No. 4971616, entitled "Process for Preparing Organic Compost from Municipal Refuse" was awarded to Mark E. Glogowske on November 20, 1990 (17). The patent involved the use of earthworms for treatment and disposal of shredded cellulose refuse.

The earthworm *Eisenia foetida* is known to contain bactericidal enzymes. In 1990, Amaravadi et al. tested the earthworm for virudical activity using Cowpea mosaic virus (CPMV) and Tobacco mosaic virus (TMV) as model agents (18). Earthworms were fed cellulose saturated with a virus suspension and their excreted castings were analyzed for structurally intact virus protein using enzyme-linked immunosorbent assay (ELISA) and virus infectivity by local lesion assays. Observations of the feeding experiments indicated a considerable reduction in the infectivity of both viruses. Virucidal activity was also observed when virus suspensions were incubated with the earthworm enzyme extract and analyzed by local lesion assay. The observed reductions in the infectivity of both viruses suggested that *E. foetida* might possess a virucidal enzyme system and, accordingly, might contribute to the inactivation of pathogenic viruses potentially associated with land application of sewage sludges and livestock manure.

Another U.S. Patent No. 5055402, entitled, "Removal of Metal Ions with Immobilized Metal Ion-binding Microorganisms" was awarded to Greene et al. on October 8, 1991 (19). The inventors cited the use of earthworms.

3. PROBLEMS AND TECHNOLOGY BREAKTHROUGH

3.1. Introduction

While vermicomposting has demonstrated its benefits, the process faces obstacles in meeting U.S. regulatory requirements. This section presents the problems and progress made in Vermicomposting, i.e. new technologies that have been developed to overcome the technical and legal problems.

3.2. Problems

Scientific interest in earthworms is on the rise worldwide (20–26). At the Fifth International Symposium on Earthworm Ecology in 1994, 183 presentations were given at the 1994 International Symposium that were divided into two general categories: using earthworms directly in horticulture and agriculture to enhance crop growth; and using earthworms to turn various residuals into beneficial composts for reuse. Despite the increasing number of studies, however, financial support for Vermicomposting research has been cut by the funding agencies in the U.S. since 1990.

Another problem is the process's failure to meet regulatory requirements. The U.S. Environmental Protection Agency's "Process to Further Reduce Pathogens (PFRP) Requirements" for in-vessel or aerated static pile composting of biosolids requires maintaining a temperature of 55°C or higher in composting for three days. Worms can survive in thermophilic composting windrows, but they tend to stick to the edges of the pile. Temperatures above

35°C, which is the heat generated by thermophilic composting is too high for earthworms, and will kill them. In Vermicomposting, temperatures are generally kept below 30°C. While organic substances can be effectively processed by worms at low temperature range, the US EPA's PFRP requirements cannot be met. Progress in Vermicomposting of organic substances proceeded slowly due to the above technical and legal problems.

There has been continuous debate in the State of California, U.S.A., regarding the classification and potential regulation of composting facilities. A draft of regulations released in August 1994 by the California Integrated Waste Management Board (CIWMB) excludes Vermicomposting operations from the notification and permitting that would be required of most larger facilities using conventional thermophilic composting to process yard trimmings, manure, biosolids and other organic substances (24). Under current California ruling, Vermicomposting may be considered an agricultural operation, in which vermiculture uses organics as a feed stock for raising worms in a worm farm. The advantage is that the owners and operators of the Vermicomposting facilities have free rein in process control and management, and are not subject to the state inspections. The disadvantage is that as long as Vermicomposting is not recognized as solid waste disposal process, the progress for its technology development and application will be slow.

Noting the U.S. federal requirements on PFRP, vermiculturists now precompost the organic substances in the thermophilic temperature range for pretreatment and disinfection. Worms are added to compost windrows for subsequent vermiphilic decomposition after the heat of initial thermophilic decomposition subsides. In comparison with conventional thermophilic composting as a process, the modified Vermicomposting process has a shorter processing time. With conventional thermophilic composting alone, it is difficult to produce high-quality products under six months; while with the modified Vermicomposting (i.e. thermophilic composting pretreatment plus Vermicomposting post-treatment), it is possible to create a marketable end product in one-sixth the operating time. Compared to the conventional thermophilic compost end product, vermicompost contains smaller particles and worm cocoons (meaning a free work force for the future), and has lower odor and enhanced microbial activity. According to commercial estimates, consumers would be willing to pay up to three times more for the vermicompost, or worm castings, than they'd pay for most normal thermophilic compost. Many commercial-scale breakthroughs in Vermicomposting technology have been noted and are introduced below (23–25).

The Resource Conversion Corporation (7825 Fay Avenue, Suite 380, La Jolla, CA 92037, U.S.A.) has developed a proprietary "Vermiconversion System" which significantly modifies traditional vermiculture windrow methods. Variations include sloped plastic liner beneath the windrow, reclaim water, aeration piping and a sprinkler to maintain proper temperature levels. In July 1994, Resource Conversion Corporation and Sanifill, a national landfill company, together opened Canyon Recycling outside of San Diego, which is a six-acre (note: 1 acre = 4047 square meters = 0.4046 ha) facility currently processing around 100 tons per day of brush, green material, and wood from construction and demolition operations and manure from the San Diego Zoo. After grinding and screening, some woody materials are marketed "as is." Leafy greens, wood fines and manures proceed through a blending plant, then "cured" via thermophilic composting to neutralize pathogens. After curing, the preprocessed material

is applied to the vermiculture windrows in thin layers. The rows are carefully segregated and check for biological reactions to new feedstock. Two to four inches of material are applied every other day, continuously. The rows are compartmentalized to prevent possible contamination of the entire facility. The facility adopts both the thermophilic composting pretreatment (for 3 to 15 days aiming at pathogen reduction and decomposition) and the Vermicomposting post-treatment (for additional 15 to 30 days aiming at final curing and decomposition). Their worm castings product is being sold for $33 per ton on the bulk market. The company is now building a 100-acre facility to manage San Diego's biosolids under a 20-year contract.

The Oregon Soil Corporation has developed a technology to reduce the space requirements for a vermiculture operation using a "continuous flow system." The newly developed continuous flow system utilizes a raised, 120-foot trough (note: 1 foot = 0.3048 meter) that is 2.5 feet deep and 8 feet wide, with a mesh floor. An adapted manure spreader makes a daily pass over the trough, laying down about three inches (note: 1 inch = 2.54 centimeters) of prepared organic materials, or roughly six tons per day (note: 1 ton = 2000 pounds; 1 pound = 0.454 kilogram). As the worms eat up through it, the worm castings sink down, and are mechanically scraped off the bottom of the screen and collected. Under the protection of a greenhouse like structure, the worm reactor can handle about 2,500 tons of organic residuals a year. Currently, the Oregon Soil Corporation accepts year trimmings deliveries from local landscapers and picks up food scraps and paper from 15 Fred Meyers grocery stores around Portland. They process around five or six tons of food scraps, over two tons of supplemental yard trimmings or compost, and around half a ton of paper per day. It takes only 21 days to make earthworm castings using the continuous flow system.

The Worm Concern (Note: It is The Worm Connection now in California, U.S.A.) had grown to a 22-acre spread during its 18 years in business. Around 100 tons per day of brush, leaves, tree limbs, grass clippings and horse manure are delivered to the site for processing. Incoming material first passes through a grinder and a trammel before being placed in windrows by a front-end loader. The facility adopts both anaerobic windrow preprocessing (in which the piles are not turned at all until material is moved to the worm rows) and Vermicomposting post-treatment using worms. At harvest time, worm rows are scooped up with a front-end loader and placed in screen. Castings come out one end and the worms come out the other, unharmed. Their vermicastings are sold in bulk, blended on site with mulch or other landscape products, bagged for retail sale.

Finally, the Environmental Earthworm Projects, Inc. (8114 Port Said Street, Orlando, FL 32813, U.S.A.) currently operates two sites handling a combined total of 30 tons per month of composted yard trimmings from the Orange County landfill and 20 tons per month of shredded cardboard. They also have conducted earthworm trials with RDF fines from Palm Beach County and other organics.

3.3. Progress in Vermicomposting outside the U.S.A.

Engineers and scientists in the countries other than the U.S.A. have shown their interest in the theories, principles and applications of vermistabilization process since 1992. Practical applications of Vermicomposting process in disposal of biosolids and organic solid wastes

have been attempted by many entrepreneurs around the world. The progress in Vermicomposting process development and applications outside of the United States is discussed below (20–26).

In November 1992, Concheri et al. of Italy reported humification of organic waste materials during earthworm composting (20). In March 1993, Anton et al. of Spanish Council for Scientific Research, Madrid, Spain, reported carbofuran acute toxicity to *Eisenia foetida* earthworms (21).

In 1993, Van-Gestel and Ma of the National Institute of Public Health and Environmental Protection, Bilthoven, Netherlands, reported their results on development of QSAR's in soil ecotoxicology (22). QSAR stands for the "quantitative structure-activity relationship." The earthworm toxicity and its soil sorption of chlorophenols, chlorobenzenes and chloroanilines were documented by the investigators of Netherlands.

Also in 1993, Original Vermitech Systems, Ltd (2328 Queen Street East, Toronto, Ontario M4E1G9, Canada; Tel. No. 416-693-1027) installed a composting unit with a capacity of up to 600 pounds of organics per day at the Brockville Psychiatric Hospital in Ontario, Canada. It is the largest composter in Canada right now (23). The system is equipped with panels and temperature sensors for maintaining a tolerable environment for the worms.

At the Fifth International Symposium on Earthworm Ecology, held at Ohio State University in 1994, scientists from the University of Agricultural Sciences in Dharwad, India, told conference attendees that in their experiments, earthworms could turn crop and weed residuals into vermicompost at the rate of 8 to 10 tons per year from a bed area of 100 square meters. (24, 25). At the same symposium, scientists from the Biosystems Research Group at the Open University, Milton Keynes, in England, reported on their experiments of the modified vermicomposting process (24, 25). The English scientists added earthworms to compost windrows after the heat of initial decomposition subsided. Their worms worked well in this situation and shortened the time of curing and stabilization of the compost.

Changes in heavy metal extractability and organic matter fractions after Vermicomposting of sludges from a paper mill industry and wastewater treatment plant was reported by Elvira et al. of the University of Vigo, Spain in 1995 (26). According to the researchers from the Department of Natural Resources, University of Vigo, Vermicomposting of paper mill sludge has been proven to be viable in their country.

4. PIONEERS, CURRENT STATUS AND RESOURCES

The pioneers of the vermistabilization process, as well its current status and resources, are introduced in this section in detail.

4.1. Pioneers and Current Status

Many pioneers of Vermicomposting process deserved to be recognized. Jack E. Collier and Diane Livingstone were principal investigators of a milestone research project sponsored by the National Science Foundation entitled "Conversion of Municipal Wastewater Treatment Plant Residual Sludges into Earthworm Castings for Use as Topsoil" (7). Collier and his wife still operate an earthworm farm in California, U.S., which provides high-quality earthworms

for all types of earthworm research including vermistabilization. The Colliers often serve as consultants on their vermistabilization technology to individuals or organizations. Dr. Mark Buchannon, a soil scientist of the University California at Santa Cruz, U.S., recently collaborated with the Colliers to complete his Ph.D. research in a similar field.

Dr. Raymond C. Loehr of the University of Texas at Austin, Department of Civil Engineering, Austin, TX, U.S., is another legend in vermistabilization technology development (11–13). Dr. Loehr, too, consults on vermistabilization research and applications, if requested.

Dr. Clive Edwards, Professor of Entomology at Ohio State University, has also been instrumental as the founder of the International Symposium on Earthworm Ecology, and has conducted several key Vermicomposting projects leading to commercialization of the process.

Practicing Vermicomposting technologists who can provide assistance in vermicomposting facility installation and process operation include: Frank Stevenson of the Environmental Earthworm Projects, Inc., Dan Holcombe of Oregon Soil Corporation, Albert Eggen of Original Vermitech Systems, Ltd., Joseph Roberts of Resource Conversion Corporation, Tim Morhar of The Worm Connection, and Sandra Kandracs of Enviro-Ganics.

Writers/reporters Gene Logsdon, David Riggle and Hannah Holmes discussed the progress of vermicomposting technology in two articles for *BioCycle* (24, 25), a trade journal that documents and reports the scientific knowledge and commercial news involving worms.

Steven Zorba Frankel and Stephen White of the Edible City Resource Center, has published a 32-page quarterly newspaper, *Worm Digest* (27–38), which promotes Vermicomposting technology as well as other technologies involving the use of earthworms. Today *Worm Digest* reports on the subjects of worms and worm composting for organic waste conversion and soil enrichment. The newspaper generally features a wide variety of interesting and practical information to help promote awareness of vermiculture eco-technology on all levels. Columns such as the following appear intermittently in each issue (27–29): Worm Shorts × New Products × International Worm News × The Industrious Worm (large-scale projects) × Hands-On × Worm Workers × Kids' Corner/Page × Questions & Answers × Eco-Logic × Worm Stories × Cyber-Worm × Advertisements & Resource Listings × Calendar of Events.

At the request of environmental engineers in Ukraine, the authors conducted an investigation on the current status and future direction of vermistabilization process. It was discovered that the vermistabilization (Vermicomposting) operations/research in sites such as Syracuse, NY, Ithaca, NY, West Chester, PA, San Jose, CA, and Austin, TX in the U.S. was terminated due to minor technical and legal problems, and a lack of financial and public support. It is encouraging to learn, however, that several companies in the United States and Canada have seriously conducted their research for modification and optimization of the Vermicomposting (or vermistabilization) process despite the lack of proper funding. Now the process has been improved and commercialized, and many large-scale Vermicomposting or vermiculture projects in Florida, California, Oregon and Ontario are in progress.

Earthworm research is still being widely conducted by soil scientists and environmental scientists around the world. Earthworms are tested as the organisms for organic waste disposal, the toxicity indicators of ecological system, or as the topsoil producers. As mentioned, there is even an annual International Symposium on Earthworm Ecology.

Interest in vermistabilization process for biosolids management has quickly spread from the U.S. to European countries (20–26), indicating that there will always be ample room for additional research on process improvement.

To explore or establish any international cooperative programs in the field of environmental engineering, readers are encouraged to contact the authors and the experts listed in Section 4.2 for technical or managerial assistance. Important resources of vermicomposting process around the world are introduced elsewhere (40–66).

4.2. Resources

Important resources of Vermicomposting process around the world are introduced in this section. It should be noted that the first letter of each resource defines its nature in accordance with the following *KEY: Associations (A), Publications (P), Retail Businesses (R), Consultants (C), Distributors (D)*

P 1. Edible City Resource Center, Worm Digest, PO Box 544, Eugene, OR 97440, U.S.A.
R 2. The Worm Factory, RR # 3, Perth, Ontario, Canada K7H 3C5.
A 3. The Composting Council of Canada, Canada. e-mail address: ccc@compost.org.
A 4. Association of Oregon Recyclers, PO Box 15279, Portland, OR 97210, U.S.A.
P 5. BioCycle, Journal of Composting & Recycling (monthly) 419 State Avenue, Emmaus, PA 18049, U.S.A.
W 6. Lake County Worm Farm, PO Box 1332, Kelseyville, CA 95451, U.S.A.
P 7. Australian Worm Growers Association, PO Box 318, Ferntree Gully, VIC 3156, Australia.
R 8. Arlan & Sons, (bookseller) 11881 Arroyo, Santa Ana, CA 92705, U.S.A. e-mail address: arlan@neptune.net.
R 9. Avant Garden Vermicomposting Systems, (worm bins) PO Box 1047, Point Reyes Station, CA 94956, U.S.A.
C 10. Vermitechnology Unlimited, Inc., PO Box 130, Orange Lake, FL 32681, U.S.A.
A 11. International Worm Growers Association, PO Box 887, Littlerock, CA 93543, U.S.A.
R 12. WormWide Books, 20 Forest Avenue, Kingston Park, South Australia 5049, Australia.
W 13. Rainbow Worm Farm, 24700 County Road, No. 95, Davis, CA 95616, U.S.A.
C 14. Oregon Soil Corporation, 1324 Beaver Lane, Oregon City, OR 97045, U.S.A.
R 15. Flowerfield Enterprises, 10332 Shaver Road, Kalamazoo, MI 49002, U.S.A.
C 16. Roberta Trombley, 3030 Marshall, Cincinnati, OH 45220, U.S.A.
D 17. Viscor Distribution Inc. (Worm Bins), 12165 Cherrywood Drive, Maple Ridge, BC, Canada V2X OB7.
R 18. Worms & Worm Boxes, 968 Valencia Street, San Francisco, CA 94110, U.S.A.
W 19. Willingham Worm Farm, Rt. # 1, Box 241, Butler, GA 31006, U.S.A.
W 20. Manchester Worm Farm, 1131-0 Tolland Turnpike, Manchester, CT 06040, U.S.A.
C 21. Environmental Recycling Systems, PO Box 904, Alpine, CA 91903, U.S.A.
C 22. Vermiculture Services International, U.S.A.
D 23. Recycle-It Corporation, U.S. (distributor of worm bins, curbside recycling bins and backyard composting bins) Tel. No. 800-769-1044.
W 24. Olympic Worm Casting Farm, McCleary, WA, U.S.A.
C 25. Casting a New Future, Portland, OR, U.S.A.
D 26. RPM, 2829 152ND Ave. NE, Redmond WA 98052, U.S.A.
27. Sound Resource Management Group, Inc., 119 Pine Street, Seattle, WA 98101, U.S.A.
R 28. Worm World, 26 Ihnat Lane, Avella, PA 15312, U.S.A.

C 29. Resource Conversion Corporation, 7825 Fay Avenue, Suite 380, La Jolla, CA 92037, U.S.A.
C 30. Environmental Earthworm Projects, Inc., 8114 Port Said Street, Orlando, FL 32813, U.S.A.
C 31. Original Vermitech Systems, Ltd., 2328 Queen Street East, Toronto, Ontario, M4E1G9, Canada.
C 32. The Worm Connection, 581 Camino Manzanas, Thousand Oaks, CA 91360, U.S.A.
C 33. Zorex Corporation, PO Box 405, Newtonville, NY, U.S.A, 12128-0405.

5. PROCESS DESIGN CONSIDERATIONS

5.1. Process Adoption and Advantages

Earthworm castings are essentially odorless when dry; when damp, they have a mild odor like a good quality topsoil. Also, castings have a favorable appearance. When sifted and dry, they are granular, about 0.02 to 0.1 inches (0.5 to 3 mm) in maximum dimension (with some fines); color is brownish gray. In a study where municipal sludge was applied to a wheat crop, it was found that when earthworms were added to the sludge, the germination rate of the wheat improved (50). The odor, appearance, and soil supplementation advantages of the earthworm conversion process may help in the acceptance of biosolids by farmers and householders.

Earthworm conversion affects several other biosolids characteristics. The oxygen uptake rate increases (46); the acid-extractable fraction of various nutrients increases (50). The volatile content of the solids drops slightly and humic acid concentrations fluctuate (46). While these effects may be beneficial, there are no data to show how the results affect design or operation of earthworm conversion installations.

The earthworm conversion process would appear to be low in cost, although this cannot be said with certainty, since no cost data are available for full-scale operations on biosolids. The process does not require chemicals, high temperatures, or large amounts of electricity. Only a small amount of low-speed mechanical equipment is needed. Significant expenditures may be required to offset the potential operating difficulties discussed below.

5.2. Process Operation and Troubleshooting

A number of potential operating difficulties and their solutions exist in the earthworm conversion process. None of these difficulties are insurmountable, however. Probably the most difficult problem is to economically pretreat anaerobically digested biosolids so that it is nontoxic to the worms (59, 62). Other problems that must be considered include:

Worm drowning: Worms must be protected from flooding.

Worm loss due to migration from the process: Caused by flooding, toxic sludge, unpalatable sludge, adjoining areas attractive to worms, lack of artificial lighting on rainy nights.

Toxicity of sludge to worms: Significant for anaerobically digested sludge. However, toxicity is eliminated by exposing the sludge to air for two months (46) or wetting sun-dried sludge daily for 14 days (50). Stabilization by lime or chlorine is not recommended for sludge that will be fed to earthworms. Toxicants such as copper salts might also cause problems. Aerobic digestion is best suited for sludge to be converted by earthworms.

Toxicity or unpalatable nature of dewatering chemicals: Avoided at Hagerstown, Md., by use of food-grade polymer (48). Drying beds may be used; drying beds do not usually require chemicals.

Worm shortage in the process, so that worm additions are required: Worms reproduce via egg capsules, which may be lost from the process in the castings. Also, toxic conditions, drowning, and other problems will cause worm populations to drop. At Hagerstown, Md., a worm raising operation has been proposed to supply the necessary make-up worms to the sludge conversion process (48).

Shortage of worms for initial inventory or restart: To begin operation, a large worm inventory may be needed, but local worm suppliers may be unable to meet this demand. Gradual start-up is therefore desirable, especially for large plants. Also, earthworm exchanges may become available nationwide so that sludge operations can draw on larger numbers of earthworm suppliers.

Temperature extremes: Worm feed most rapidly at 15 to 20°C; about 5°C, feeding is quite slow (46). Freezing will kill worms. High temperatures can also cause problems. It may be necessary to stockpile sludge during the winter or provide a heated building for the conversion process.

Shortage of enzymes: Not a problem, despite claims by marketers of enzyme preparations that these preparations are valuable to the process (52).

Exposure to light: Worms avoid bright light. Some sort of cover or shade should be provided so that worms will convert the top layer of the sludge.

Dehydration: There is a minimum moisture content for the worm bed (52).

Salinity in castings: Under some conditions, castings may have sufficient dissolved salts to inhibit plant growth. This problem may be eliminated by leaching or by mixing the castings with other materials with lower dissolved salts (53, 54).

Contamination of castings by heavy metals, motor oil, rags, and similar materials: Source control may be used where feasible, as for other processes aimed at reuse of biosolids as a soil conditioner.

Odors: The most likely source is raw or aerobically digested sludge, which has been stockpiled to await earthworm conversion.

5.3. Process Limitations

Limitations of the earthworm conversion process include, but are not limited to, the following (58, 59):

Earthworm conversion decreases the total nitrogen values in the biosolids, as ammonia nitrogen will be lost to the atmosphere.

Costs are unpredictable.

Two common ions in municipal wastewater biosolids, ammonium and copper, may be toxic to worms. Studies have found that these ions were lethal at additions equivalent to 180 mg NH_4^+-N and 2,500 mg Cu per kilogram of wet substrate (55, 56). Safe limits for these elements are not known.

Cadmium accumulates in the worm Eisenia foetida. Zinc apparently does not accumulate in Eisenia foetida but does accumulate in other species (56, 57). If the worms are to be used as animal feed, the system must be operated such that cadmium and zinc concentrations in the worms do not exceed recommended levels for animal consumption.

Space requirements may rule out earthworm conversion at some treatment plants.

The earthworm business has been afflicted with unsound investments and excessive claims. For example, it has been claimed that earthworm processing is able to reduce concentrations of heavy

metals (58). Any such reduction could only be caused by simple dilution with uncontaminated waste or by concentration of the contaminants in the earthworms.

If a particular sludge is suitable for earthworm conversion, that sludge should also be suitable for reuse as a soil conditioner without being processed by earthworms. However, earthworm conversion reduces odor, improves texture, and may increase germination rate.

These limitations may seem significant but are not overwhelming. Considerable research and development is underway, and it appears that earthworm conversion may soon have a role in municipal wastewater treatment plant sludge processing.

5.4. Process Design Criteria

Design criteria have been generated by the operators in the field (46–49, 59, 61) for the vermicomposting process.

Species of worm being tested were *Eisenia foetida* (redworm, hybrid redworm, tiger worm, dung worm) (46, 49), *Lumbricus rubellus* (red manure worm, red wiggler worm) (47), and *Lumbricus terrestris* (nightcrawler) (46). The following are the compiled design criteria:

Detention time of sludge in worm beds = 2 to 32 days (47, 48)
Worm reproductive cycle = 1 to 2 months
Rate of worm feeding (15°C) = 0.17 to 1.7 grams dry sludge per gram dry worm weight per day (46)
Optimum temperature = 15 to 20°C
Dry matter content of worms = 20 to 25 percent (*Eisenia foetida*) (49)

Minimum solids content of the worm bed mixture = 20 percent; Actual minimum solids content depends on such factors as porosity, type of sludge, ability to keep aerobic. Experiments are being conducted to better define these parameters.

6. PROCESS APPLICATION EXAMPLES

The Wright-Patterson Air Force Base in Dayton, Ohio, U.S.A., (43) launched a vermicomposting program in July 2002, using earthworms to consume a daily average of 500 pounds of solid waste. The worms digest vegetable matter and old newspapers, saving the base about $25 per day on transporting and disposing of waste. As the number of worms grows, so does the amount of waste they consume. The base acquired 250,000 worms and their climate-controlled home (at a constant 70 degree temperature) for the environmental project. At the base, which produces fruit and vegetable waste from its commissary, the earthworms have flourished, now numbering more than 300,000. Their numbers eventually could top 1 million. The worm casings replace chemical fertilizer at the base's golf course, which saves additional money. More successful stories can be found in the literature (40–66).

Vermicomposting has gained popularity in schools and municipalities, according to Stuckey and Hudak (60). In Boston, Massachusetts, U.S.A., Josiah Quincy Elementary School received a grant to build a roof top organic garden. The students maintain garbage-eating red wiggler worms to break down fruits and vegetables. Once processed in the bin, the compost is applied to the garden. In Orange County, Florida, U.S.A., a revolutionary worm-use

concept has been promoted where worms stabilize biosolids to a "Class A pathogen standard" substance.

7. FUTURE DEVELOPMENT AND DIRECTION

Vermicomposting (or vermistabilization) should be encouraged by governments in the field of environmental engineering as a promising process for disposal of biosolids and other organic solid wastes (65). Special efforts should be made in the near future to obtain recognition for the process, and funding sources should be explored at all levels for economical analysis and optimization of the process. At the global level, international agencies should encourage and fund the transfer of Vermicomposting technology between the U.S.A. and other countries (66).

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18
Biological Odor and VOC Control Process

Gregory T. Kleinheinz and Phillip C. Wright

CONTENTS

Abstract Most of volatile organic compounds (VOC) and odor-causing organic substances in air streams are biodegradable, thus can be effectively removed by a gas-phase biological filter or a biological scrubber. The topics covered in this chapter include: odor classification, odor emission regulations, odor control technologies, biofilters, biological scrubbers, process control, monitoring, and biofiltration case studies.

Key Words gas-phase biofiltration • biological scrubber • gas-phase attached growth biological process • air pollution • odor removal • VOC removal • volatile organic compounds • process design • case studies.

1. INTRODUCTION

Biofiltration is the use of microorganisms, immobilized on a biologically active solid support, to treat chemicals in an airstream. While the term implies a physical process, the process is biochemical and will not likely be changed in the near future. Biofilters have been used for volatile organic compound (VOC) abatement, mitigation of odor-causing compounds, and in conjunction with other treatment technologies (i.e., soil vapor extraction). With recent

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Table 18.1
Industries that have used biofiltration for air pollution control

Animal Facilities, Large-scale	Painting Operations, Large-scale
Automotive	Petrochemical Manufacturing
Chemical Manufacturing	Petroleum
Coatings	Plastics Manufacturing
Composting	Printing
Ethanol Production	Pulp and Paper
Food Processing	Rendering
Fragrance	Semi-Conductor
Iron Foundries	Sewerage Treatment
Landfill Gas Extraction	Wood Products

changes in U.S. air regulations, increased pressure has been placed on industries that emit chemicals into the air. Biofilters have been an increasingly popular choice as a treatment option due to their low operating cost and relatively low capital costs, when compared to other technologies. Biofilters operate under the premise that contaminants in the airstream partition into an aqueous layer on the solid support, where it is bioavailable, and then degraded by the microbial community present. Complete metabolism of an organic compound yields carbon dioxide and water, which is then moved out of the biofilter. In general, conventional biofilters have been the most successful in applications with low flow rates, and relatively low concentration of contaminants. Table 18.1 illustrates some of the industries that have used biofilters.

Biological treatment methods have been widely used by industry to mitigate environmental contamination throughout the twentieth century. However, only recently has biofiltration gained acceptance in the United States as a viable treatment alternative for air emissions. Some of the impetus for this adoption was the Clean Air Act Amendments (CAAA) that were put into place in 1990. This brought air emissions into the forefront of legislative and regulatory agencies throughout the U.S. In addition to formal regulations, a lack of tolerance has been seen in recent years for unregulated odorous emissions. These types of odorous emissions are typical of wastewater treatment facilities and are largely unregulated. Prior to their adoption in the U.S., biofilters had enjoyed much success in Europe, particularly the Netherlands, as a viable treatment alternative to a variety of air emission issues. In fact, the first biofilters are rumored to date back several hundred years to the mitigation of odors from outdoor privys; the first U.S. patent was granted in 1957 to Pomeroy (5). However, early systems often used porous soil materials as a solid support, and primitive piping systems for airflow through the beds. These first attempts were moderately effective, but were prone to channeling and poor air distribution.

Biofiltration has come a long way since 1957 and the market is expected to increase in the future. It has been estimated that the biofiltration industry would be over $100 million dollars in 2000 (1). To our knowledge these numbers have not been verified, although biofiltration companies in the United States have seen unprecedented growth over the past five years.

Given the comparable capital costs of biofilters and low operating costs relative to competing technologies, it is likely the market will continue to grow and evolve. For example, many wastewater treatment facilities were constructed in the 1970s, and were placed at the outskirts of their respective communities. Due to urban growth and development, residential housing, offices and businesses now surround these once semi-rural locations. With increased exposure to populations of people, wastewater treatment facilities are under increasing pressure to mitigate odors on site. Since it is usually not practical to move the facility, odor-control technologies must be implemented on site. It should also be noted that the wastewater industry is not the only industry being impacted by decreased tolerance for odorous air emissions.

2. TYPES OF BIOLOGICAL AIR TREATMENT SYSTEMS

2.1. General Descriptions

In conventional packed bed biofilters (Figure 18.1), the vessel contains a layer, often 1 to 1.5 meters thick, of some type of filter material such as compost or peat. The waste gas, which is usually pre-humidified to help prevent bed dry-out, percolates up through this packed bed. Water sprays, or drip feeds (Figure 18.1), are positioned over/in the bed to add extra moisture to also prevent dry-out, provide a source for pH control, or to supply additional nutrients. The bed is run in a minimum liquid condition to reduce pressure drop, avoid wastage, and reduce entrainment of bacteria and production of anaerobic zones. That is, the interparticle space is largely air and the water phase is stationary on the surface of the solid support. Microorganisms are fixed within a biofilm on the solid support in this type of application. Airflow may be either up-flow or down-flow depending on the engineering at the site, and

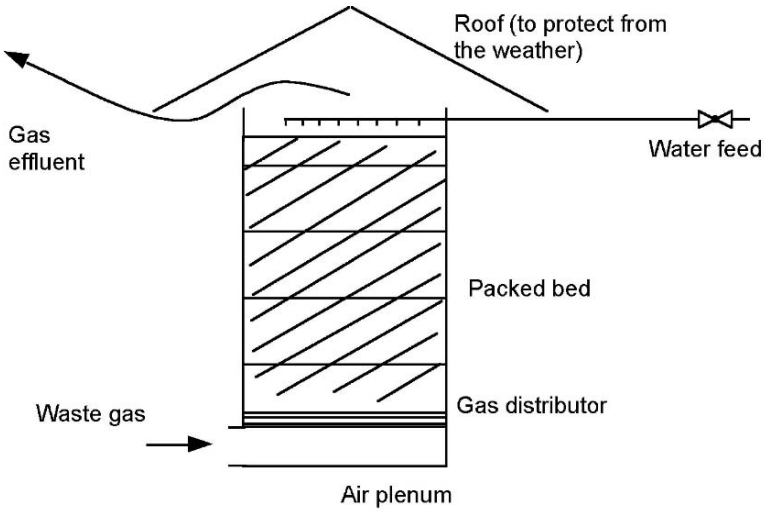

Fig. 18.1. Schematic diagram of conventional packed bed biofilter.

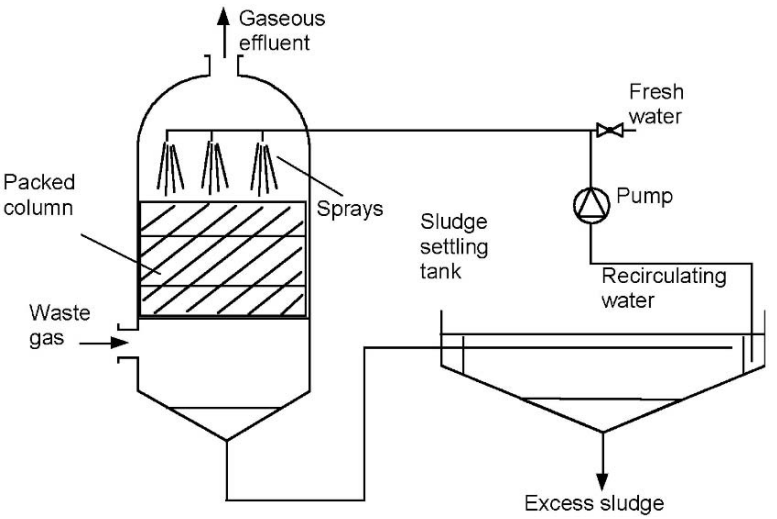

Fig. 18.2. Schematic diagram of conventional trickle bed biofilter.

results of pilot studies. Both airflow directions have demonstrated successes and failures, with other factors being more critical to the success of the system.

Trickle bed reactors differ from conventional packed bed biofilters in that the packing material is often synthetic packing (Figure 18.2), such as tellerettes or pall rings, and the liquid feed into the column is much greater. The liquid phase, after trickling through the column, passes into another tank to allow settling of solids and additional biodegradation before being pumped back. The interparticle space is largely water-filled, with the water-phase flowing through the media. Microorganisms are fixed within a biofilm on the solid support in this type of application. Airflow in this type of system is usually up-flow, or counter-current to the water flow.

The third treatment method, bioscrubbing, involves absorption of the target species into a liquid that is sprayed counter-currently to the gas flow in a tower (Figure 18.3) contactor. The liquid phase containing the target species is then pumped around to an activated sludge tank (Figure 18.3) where the biodegradation occurs by using freely suspended microbes. The liquid phase is then pumped back to the absorber tower's spray feed system.

A large number of disadvantages (summarized in Table 18.2) prevent the widespread development of biological waste gas abatement, despite its advantages (see Table 18.6) (2–6).

2.2. *Novel or Emerging Designs*

As discussed previously, biofilters have several limitations. Among these limitations are problems with high influent concentrations and toxicity, high flow rates and low retention times, and low solubility compounds with poor degradation. Generally, biofilters have been applied to airstreams containing high flows with low levels of contaminants. Also, biofilters

Table 18.2
Disadvantages of conventional biological odor abatement technologies

Conventional biofilters	Conventional trickle bed biofilters	Conventional bioscrubbers
• Packing is usually non-homogeneous, often preventing a uniform gas distribution → short-circuiting. • Low specific gas flow (average for compost beds 150 m^3-gas/hr/m^2-bed, max up to 500 m^3-gas/hr/m^2-bed). • Aging phenomenon, resulting in: • Lumping. • Drying out. • Developing of anaerobic zones due to moisture accumulation. • Development of shrink cracks. • Bed compaction. • Difficulty in maintaining an even bed pH.	• Biological overgrowth leading to increased Δp_{bed}. • Low specific area to reduce Δp_{bed}. • Drain water has to be continuously separated from excess biomass before being recycled. • Fresh water must be constantly fed to the system due to losses. • Non-homogeneous temperature and concentration profiles.	• More energy intensive than conventional packed bed biofiltration. • Due to large amount of liquid there is a danger of active microorganisms being carried away. • More sensitive than packed bed biofilters to feed fluctuations. • Operation takes place in more than one unit. The sludge tank often requires extra stirring and oxygenation. • Periodic removal of sludge.

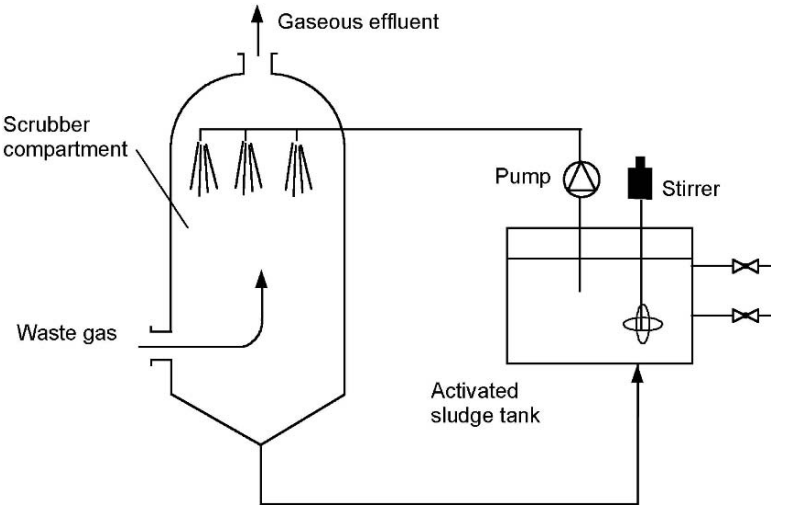

Fig. 18.3. Schematic diagram of a conventional bioscrubber.

have been traditionally applied to situations where the airstream contains relatively soluble compounds. However, since these limiting factors have been long realized, there has been substantial development of technologies that address these limitations, so that biofilters may be used in a wider range of applications. If gas-phase biofiltration is going to receive increased

take-up industrially, it is vital that the stability, efficiency and range of operating conditions are improved. This section briefly addresses a number of the potential emerging technologies.

2.2.1. Pollutant solubility in the aqueous phase

As described earlier, pollutant solubility may be an issue dictating method choice. Some researchers have attempted to address the limitations of water solubility by using surfactants in the biofiltration beds (7, 8). The theory is that the increased solubility of the chemicals in the bed will increase partitioning into the liquid phase and thus make the chemicals more bioavailable. Lab studies have met with some success in applications dealing with chemicals produced by the forest products industry. In addition to attempts to increase solubility, changing the airflow rate has also been attempted. Since the process generally sets the airflow rates, the changes have to be made prior to the biofiltration system. These changes in flow rate are accomplished via adsorption/desorption systems. That is, high flow, low concentrations are adsorbed on to a suitable substrate (i.e., activated carbon) and are then desorbed at a lower flow rate and possibly higher concentration (9). By lowering airflow rates, the biofilter has more contact time with the chemicals and high degradation rates. The total loading on the system can be precisely manipulated to achieve the highest degradation possible. Several companies are currently marketing systems that operate on this principle. In addition, the automotive paint industry has used this adsorption/desorption technology to trap airborne pollutants and send high concentrations to thermal oxidizers.

2.2.2. Mobilized Bed Biofilters

As discussed in Table 18.2, conventional technologies may be limited due to mass transfer or mixing limitations. Three-phase fluidized (or mobilized) beds may be an alternative to conventional packed bed biofilter and absorber/scrubber/trickle bed methods. They have a number of inherent advantages for multiphase contacting, such as good inter-phase mixing and heat and mass transfer performance. This contactor type also removes the disadvantages of poor moisture and temperature control inherent in other vapor-phase biofiltration systems.

There are some limited studies into this area (e.g. (10)); however more work is required before their widespread use is acceptable, particularly in relation to process control and biological support matrices. Having said that, there are some industrial examples of mobilebed types of biofilters/bioscrubbers, such as the SC Bioreactor^{TM} system in the UK (Waterlink/Sutcliffe Croftshaw Ltd - Lancashire, UK).

2.2.3. Integrated/Train Processing

Some preliminary lab studies have been conducted which combine biological treatment technologies into "treatment trains" for the treatment of complex waste streams containing chemicals with very different chemical properties (11, 12). These systems combine the benefits of other reactor systems such as liquid reactors or chemical catalytic reactors (i.e. fast degradation rates, or the ability to degrade more complex species) with biofilters for the removal of highly volatile compounds such as methanol and 2-propanol. By treating systems with "treatment trains," airstreams with over 10,000–15,000 ppm_v of VOCs can be successfully treated at >95% efficiency.

As an example, one of the possibilities is to use catalytic combustion to partially deconstruct the VOC molecules. Catalytic combustion is often not suitable alone, as the by-products are often toxic in themselves. Therefore suitable downstream treatment is important, and biofiltration offers a cost-effective route (12).

2.2.4. Extremophilic systems

The operating window of many biofiltration systems is being widened by the application of so-called extremophiles, that thrive under conditions that normal microorganisms may find intolerable. For example, temperatures of over 60–80°C have been demonstrated, as have extremes of pH (both high and low), tolerance to high concentrations of pollutants and extremely high salinity.

3. OPERATIONAL CONSIDERATIONS

3.1. General Operational Considerations

In order to understand biofilter operation, we must look at some important terminology related to the operation of biofilters. The term "empty bed residence time" (*EBRT*) refers to the amount of time some unit of influent air would take to pass through the empty biofilter bed space. In general, this is expressed as:

$$EBRT = \left(\frac{V_b}{A_f}\right) \tag{1}$$

where V_b = volume of the biofilter bed (m^3 or ft^3) and A_f = airflow rate ($m^3\ hr^{-1}$ or cfm). The *EBRT* is always larger than the true residence time of the air passing through the biofiltration system. This is due to the fact that the solid support medium occupies a significant amount of the total area in the bed. The *EBRT* should not be used as a true measure of treatment time due to the highly variable nature of the solid support material. The "true bed residence time" (*TBRT*) can be expressed as:

$$TBRT = \left(\frac{V_b \times M_p}{A_f}\right) \tag{2}$$

where M_p = the medium porosity. Medium porosity can be anywhere from 20–80% depending on the intraparticle (space within individual particles) and interparticle (space between different particles) porosity. Porosity can be defined as:

$$M_p = \left(\frac{V_s}{V_{ss}}\right) \tag{3}$$

where V_s = volume of a given space and V_{ss} = volume of solid support material. The porosity of a biofiltration medium can be determined via a simple displacement experiment in a volumetric cylinder or via more sophisticated methods such as gas chromatography and the use of inert gas flow-through experiments (13).

EBRT or *TBRT* are usually analogous values and are directly related to the performance of the biofiltration unit. Industrial biofilters have *TBRTs* that can be as short as 15 seconds

and as long as over one minute (14). These times are usually a function of the design of the system relative to the concentration and formulation of the contaminants in the airstream. More recalcitrant, less water soluble, etc. compounds require longer residence times. The longer the *EBRT* or *TBRT*, the better the removal of the biofilter. However, the airflow rate at most facilities is dictated by air change rates in buildings or by the process from which the air is derived. Thus, as a designer of biofilter systems, one's only method of changing the *EBRT* or *TBRT* is to manipulate the size of the biofiltration unit. This may appear to be fairly simple, however, cost and space may not make this an easy proposition.

When evaluating the levels of contaminants to be treated, the most utilized measurement is Volumetric Mass Loading (*VL*). Volumetric Loading is defined as:

$$VL = \left(\frac{A_f \times C_I}{V_f} \right) \tag{4}$$

where C_I = the concentration of influent (gm^{-3}). Typically the range of *VL* is 10–160 $gm^{-3}\,h^{-1}$. While loading is important in assessing a biofilter's needs in terms of size, etc. the term "removal efficiency" (*RE*) is used to express the percentage of the influent chemicals removed by the system. *RE* is defined as:

$$RE = \left(\frac{C_I \times C_O}{V_f} \right) \times 100 \tag{5}$$

where C_o = the concentration of the effluent (gm^{-3}). The term "elimination capacity" is utilized to express the overall effectiveness of the biofiltration unit and is generally expressed as:

$$EC = \frac{(C_I \times C_O) \times A_f}{V_f} \tag{6}$$

or simply as,

$$EC = (RE) \times (VL) \tag{7}$$

Elimination capacity is the best measure of overall biofilter performance, although in some instances effluent concentrations only are used for regulatory compliance. These are used for compliance purposes due to the fact that many permits are based on the total mass that may be released regardless of effectiveness of the treatment system being used. A usual necessary (legislation dictated) *RE* will be in the range of >95–99%, but at low influent loads the *REs* will be ~100%. However, as the loading increases, the *RE* will drop below 100% (see Figure 18.4). This is called the "critical load" and is used in pilot systems to help size full-size units for optimal performance. Table 18.3 lists some ECs for a variety of chemicals being treated via different biofiltration systems.

Generally, commercial biofilters will remove anywhere from 10 to 280 gm^{-3}/hr. The higher removal is typical observed in highly water-soluble and easily degraded compounds such as acetone and methanol, while lower rates are observed with more complex and less water-soluble compounds such as α-pinene.

Table 18.3
Elimination capacity values for several biofilter applications

Chemical	Maximum Elimination Capacity $gm^{-3}\,h^{-1}$	Reference
Acetone	280	12
BTEX	30	15
Hydrogen Sulfide	130	16
JP-4, Jet Fuel	65	17
Methanol	300	18
MEK	120	19
alpha-pinene	35	20
Styrene	100	21
Toluene	100	22

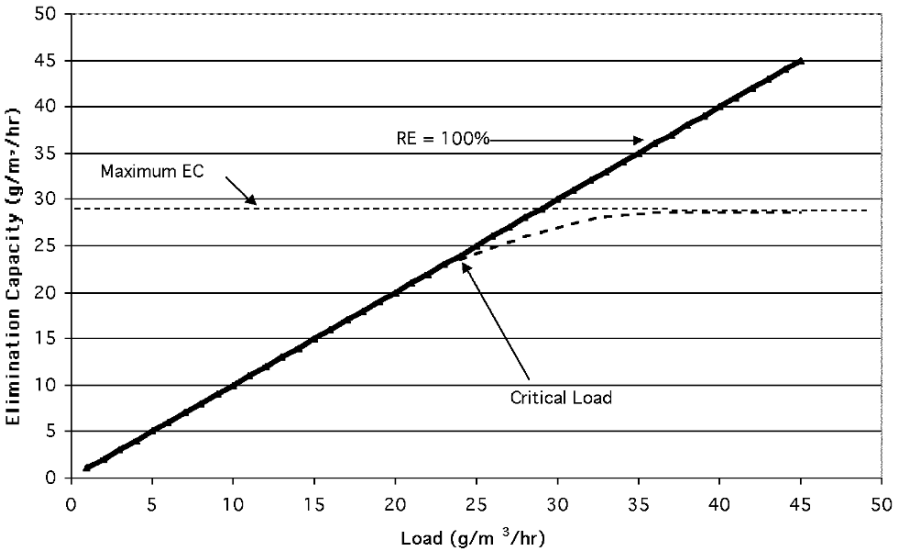

Fig. 18.4. Load versus elimination curve. The difference between the elimination capacity and the loading of the system is the RE of the system (*Source:* G. Kleinheinz, unpublished).

3.2. Biofilter Media

The choice of a solid support medium for a biofiltration system could be the most critical decision in the design of these treatment systems. Solid support media may be bioactive or inert in origin. As shall be described later, the choice will also be dictated by the type/configuration of biofilter chosen.

All good biofilter support media share several common characteristics. These include the ability to support microbial growth on the surface of the particles. Materials that have rough surfaces, significant intraparticle porosity, and no inhibitory properties to bacteria are

generally good at supporting a microbial population. The ideal situation is that they are resistant to breakdown and subsequent compaction. The breakdown and compaction of the media leads to numerous operational problems, and requires that the media be replaced more often, thus adding cost. Often materials such as perlite are added as an aid to stop bed compaction. The medium should possess adequate water-holding capacity; usually between 40–70% for bioactive media and 30–60% for some inerts. Unless the biofilter is of unique design, the media should possess a pH of between 6 and 8, and ideally would have some buffering capacity. The cost of the media relative to its lifetime should be acceptable to the operator. Each type of medium has a different cost and lifetime associated with it. It is critical that this be considered in the design, as media replacement can be a significant portion of the operating costs of a biofilter.

3.2.1. Bioactive media

Some advantages of natural biofilter media are the relatively low cost and its high availability. Natural materials such as compost and wood chips are readily available. However, they often vary significantly in their composition from one time/place to another. Bark chips can be an effective medium, but the choice of wood species is very important. For example, Douglas fir bark resists degradation more than pine bark and would save the operator the cost of media replacement and operations via lower energy costs. While natural media have several advantages, they often encounter problems with breakdown and compaction that lead to channeling and large pressure drops across the systems. Once the media starts to break down, it can lead to significant increases in operating costs due to increases in energy costs. Most importantly, the degradation of natural media can lead to poor performance of the system in terms of removal efficiency. Natural solid support media can range in price from $\$10/ft^3$ to more than $\$75/ft^3$ for such things as bagged bark.

3.2.2. Inert media

Inert media has one obvious advantage: It does not break down, as natural material will. The life is often much longer and there is little, if any, degradation of the media due to microbial activity or chemical effects in the system. This allows for long-term operation and very consistent operational parameters (i.e., flow rates, pressure drops, etc.). Nevertheless, inert media may have several disadvantages. Inert material can be much more expensive than the more readily available natural material (although not always). Furthermore, many inert materials do not have much in the way of inherent nutritional value (N and P) for supporting microbial populations, and thus rely more on the addition of these materials. Synthetic materials can range in cost from $\$40/ft^3$ for lava rock to $>\$100/ft^3$ for ceramic or plastic supports.

When a biofilter is being designed for a particular application, it is critical to evaluate the physical properties of the chemicals in that application to select the best solid support. It is not uncommon for a biofilter vendor to sell a "proprietary" media with some biofilter designs. While these media may be appropriate, they are certainly not appropriate for all applications. In fact, proprietary media are often very costly and do not perform any better than other more readily available media. In one application a proprietary media costing several hundred

Table 18.4
Summary of important properties of common biofilter materials (23–25)

Property	Compost	Peat	Soil	Activated carbon, perlite, and other inert materials	Other inert materials
Natural microorganisms population density	High	Medium-low	High	None	None
Surface area	Medium	High	Low-medium	High	High
Air permeability	Medium	High	Low	Medium-high	Very high
Assimilable nutrient content	High	Medium-high	High	None	None
Pollutant sorption capacity	Medium	Medium	Medium	Low-high[a]	None to high[c], very high[a]
Lifetime	2–4 years	2–4 years	>30 years[b]	>5 years	>15 years
Removal efficiency	Low	Low	Medium	N.A.	N.A.
Maintenance requirements	High	High	Low	N.A.	N.A.
Space requirements	Medium	Medium	High	N.A.	N.A.
Substance adaptability	Low	Low	Medium	N.A.	N.A.
Cost	Low	Low	Very low	Medium-high	Very high

[a] activated carbon.
[b] Ref. 23.
[c] Synthetics coated with activated carbon; N.A. = not reported.

dollars a ft^3 failed in 17 days, resulting in significant down-time of the biofiltration system. The bottom line is to make sure you are aware of the needs of your exact application and pick a medium that addresses your application's needs and special circumstances (if any). Further discussion and summary of solid media choice is described in Table 18.4.

3.3. Microbiological Considerations

While there are numerous engineering considerations to be aware of when designing a biofiltration system, one should always remember that these considerations would be meaningless without an active microbial population. The premise of conventional biofiltration is that a chemical passes through the biofilter bed and is transferred from the air phase to the liquid phase that surrounds the solid support materials. This liquid phase is a biofilm where the microorganisms degrade the chemical of interest. Primarily, two forces affect the flux of chemicals from the air phase to the liquid phase. These are the aqueous solubility of the chemical, and the rate of microbial metabolism in the biofilm.

Since the degradation of target compounds always occurs in the liquid phase, biofilters must maintain a hospitable environment for the microbes present in the biofilm. Generally, biofilters operate at a neutral pH of 6–8. However, some applications require low pH systems (pH of ~2), such as the use of *Thiobacillus* species to oxidize hydrogen sulfide and other reduced

sulfur compounds. At neutral pHs, numerous genera have been identified in operational biofilters including *Pseudomonas*, *Alcaligenes*, *Xanthomonas*, and several others. While these organisms have been implicated in biofilter operation, there is likely to be a consortium active in a successful biofilter working together to degrade the chemicals of interest.

It is generally accepted that many types of microorganisms contribute to the overall degradation of the chemicals in the system. This includes bacteria, protozoa, and fungi. While microbial metabolism is required for destruction of the target chemical, too much metabolism can lead to biomass overgrowth and subsequent clogging of the biofilter bed. To compound this issue, filamentous fungi can cause significant decreases in performance with only modest increases in growth due to their highly filamentous nature. Thus, when considering the growth of these systems, it is desirable to achieve a balance between chemical input, microbial growth and microbial death. The sum of this would be a constant microbial population that could be maintained consistently over a relatively long period of time.

There has been some debate regarding the effectiveness of inoculating biofiltration units with microorganisms. It is safe to say that synthetic media require some sort of microbial inoculum. However, natural media may or may not require such inoculum. The capabilities of the indigenous microorganisms should be evaluated at bench/pilot scale to determine if they possess the required metabolic capabilities. Should the necessary organisms be present, classical microbial ecology theory suggests that the microbes most adapted (fastest degraders or most capably of surviving in the system) will outcompete those less adapted. While inoculating may not harm a biofilter system, it may be a waste of time and resources. Conversely, inoculating synthetic media with specially selected microbes (from a laboratory enrichment for example) may significantly increase degradation rates. This inoculum may not grow in the system at a steady-state level and may lead to an overgrowth in the system and subsequent operational problems.

3.4. Chemical Considerations

It has been shown that malodorous gases often contain a rich "cocktail" of chemical species (4). Such typical compounds include hydrogen sulphide (H_2S), mercaptans, volatile organic and inorganic compounds (VOCs and VICs), volatile fatty acids, aromatic and aliphatic compounds and chlorinated hydrocarbons. These gases can obviously pose an environmental threat in addition to their unpleasant odor. Therefore the chemical nature of these compounds is important when choosing a biofiltration option, if possible. This section discusses the most important issues to take into account when examining the pollutant one is trying to abate.

3.4.1. Biodegradability

It has been reported that not all VOCs (4), and indeed other classes of compounds, are easily biodegradable. This results in incompatibility of the technology for all pollutant chemicals. As environmental legislation becomes tighter, more novel and efficient technologies for gas treatment will become necessary. The comparison of the relative ease of biodegradation of a number of typical pollutants is presented in Table 18.5.

A number of research challenges exist to ensure the total removal of pollutants. The "big picture" is how to modify existing bioreactors for the removal of major pollutants. The

Table 18.5
Comparison of biodegradability of various chemicals adapted from (4)

Rapidly degradable	Slowly degradable	Very slowly degradable
alcohols	hydrocarbons	tricholorethylene
aldehydes	phenols	trichlorethane
ketones	methylene chloride	carbon tetrachloride
esters	mercaptans	polyaromatic hydrocarbons (PAHs)
ethers	hydrogen sulfide	CS_2
organic acids	nitroaromatics	monoterpenes
amines		

problem, notably with recalcitrant compounds such as trichloroethylene and poly-aromatic hydrocarbons (PAHs), is that the size of the reactor to provide exit air of an approved standard is often enormous. In the rare areas of high land availability, this inefficient use of space is not a problem. Emerging technologies are being developed to solve this problem, as discussed later in this chapter.

3.4.2. Solubility

In developing design considerations for biofilters, or assessing if biofiltration is an appropriate treatment technology, there are numerous chemical considerations to be aware of. One of the most important chemical parameters is the aqueous solubility of the compound(s) of interest. Since the biodegradation in biofiltration systems occurs in an aqueous biofilm, it is critical that the chemical be able to partition into this phase. Once the chemical is in the liquid-phase it is bioavailable, but not before. Chemical structure is also an important parameter to consider since some structures are more susceptible to biodegradation than others. Microbes can degrade chemicals as very different rates (Table 18.5). For highly water-soluble compounds, the rate of biodegradation in the biofilm can be directly related to the rate of chemical movement from the air phase to the aqueous phase. For compounds that are not very water soluble, the rate of diffusion from the air phase to the liquid phase may limit biodegradation (27). It is desirable to have the rates of biodegradation, etc. be correlated to the residence time of airflow through the biofilter. That is, generally the more water-soluble the compound, the more rapidly it is degraded in the biofilter and the shorter the residence time required. Conversely, the less water-soluble compounds require longer residence times due to the limiting effect of chemical diffusion. One additional consideration is the toxicity of the chemical on the microbial flora of the biofilter. Some highly water-soluble compounds, such as ethanol, may pose problems if introduced in too high a concentration. That is, the rate of solubility into the biofilm is greater than the rate of biodegradation, causing an accumulation in the biofilm and a toxic effect on the microbes (25). This toxic effect then causes a decrease in performance and a degradation of the microbial flora in the system. However, this can be addressed in some cases by using pre-acclimated highly tolerant microbial species.

Acidity may build up in the medium due to the oxidation of compounds containing sulfide or chloride, etc., which will yield an inorganic acid. These may be removed by water flushing

at regular intervals or by using a buffering agent such as sodium hydroxide, magnesium hydroxide or calcium hydroxide, etc.

3.5. *Comparison to Competing Technologies*

As can be seen in Table 18.6, the odor control techniques can be broken down into two broad categories: (a) Physical/chemical – adsorption, absorption and catalytic combustion, and (b) Biochemical – biofiltration and bioscrubbing. When deciding on an odor control strategy, a number of factors must be considered. These factors include flow rates, type and

Table 18.6
Summary of VOC abatement technologies, adapted from (6)

Technique	Advantage	Disadvantage
Reformation of the process.	• Mostly removes the need to treat the VOC.	• Nearly always impossible to remove ALL of the offending VOC.
Absorption (scrubbing)	• Low capital cost. • Reasonably high efficiency. • Method is economic at high airflow rates. • Good also for trapping particulates.	• High operating cost. • Poor performance at unsteady-state and relatively low pollutant concentrations.
Adsorption	• Relatively high efficiency esp. for hydrocarbon-based systems. • Compounds are recoverable.	• High capital cost, esp. if the unit is regenerable. • Often large units required. • Cost vs. efficiency works best for narrow operating ranges. • Prior removal of dusts and mists.
Incineration (non-catalytic)	• Reliable. • Good for varying concentrations and types of VOCs.	• Very high capital cost. • Unwanted by-products (often toxic themselves)
Incineration (catalytic)	• Lower temperatures and higher efficiency than conventional incineration process.	• Very high capital cost. • Unwanted by-products (often toxic themselves)
Masking agents	• Low capital and operating costs.	• Don't remove VOC, simply "hide" it. • Very specific. • Unreliable (no adsorption).
Dilution and dispersion	• Cheap.	• Non-positive control. • Not a removal technique.
Biological methods (e.g. biofiltration and bioscrubbing)	• Proven technology. • Low operating cost. • Good performance at low concentration of pollutant.	• Variation in efficiency depending on pollutant. • Not flexible to changes in gas stream concentration and loading. • Poor performance at high loadings, or with complex organic materials.

concentration of malodorous compounds, level of particulate matter, and stability of flows and concentrations. A decision also can be made based on comparing the lifetime costs of various treatment processes. As indicated above, biofiltration is an established technique offering the advantages of high efficiency with generally low operational and capital costs. The technology is based on utilization of immobilized bacteria or fungi in a conventional packed bed reactor. The operation relies on absorption of the vapor-phase pollutant into a wet biofilm surrounding the solid media. Subsequently biocatalytic oxidation takes place by means of the immobilized microbial species.

4. DESIGN CONSIDERATIONS/PARAMETERS

4.1. Pre-design

It is important firstly to examine the pollutant gas to be treated. Important parameters that need to be assessed are the compounds that are present in the gas stream and their concentrations. Secondly the volumetric or mass flow rate and temperature of the gas stream to be treated is required. Ideally it is of great use in the design process if one can obtain a history or a quantitative prediction of how these variables will vary, both temporally and particularly for the constituents, how much the relative concentrations will vary, and if any other compounds are likely to be present. If at all possible it is ideal, if a bench scale and/or a pilot scale study could be undertaken, to obtain a relationship between the volumetric pollutant loading (usually expressed as g $m^{-3}_{gas}\,h^{-1}$) and the bed elimination capacity (*EC*, expressed in g$m^{-3}_{bed}\,h^{-1}$). A balance is required between the *EC* and the actual amount of pollutant removed. Often regulations state that a certain percentage of pollutant must be removed rather than an actual *EC*.

From the point of view of mineralization of the pollutant, the kinetics of such a process are likely to follow an inhibition-type model form. These types of models are unstructured kinetic models generally developed, or extended, from the Monod-equation for substrate uptake (e.g. Haldane/Andrews, Levenspiel). The influence of the inhibition term becomes more pronounced as the concentration of pollutant rises (see also Figure 18.1).

Depending on the results of the study it may be important to multi-stage the treatment process. This can be because, during the biodegradation process, some of the primary compounds or their degradation products may be recalcitrant. In this way it may be possible to obtain, for example, a high *EC* for one compound, with 99+% removal, yet still be faced with ∼100% of another compound or metabolic intermediate. Intermediates can often be as environmentally dangerous as the primary compounds. So, to treat these other species, it may be economically (both from a capital cost and running cost point of view) or operationally attractive to have different stages, or even separate biofilters/bioscrubbers in the process.

4.2. Packing

Depending on the exact pollution application and bioreactor configuration (biofilter vs. bioscrubber, vs. biotricking filter) a choice as to the appropriate packing material will need to be made (see Table 18.4). However, despite a number of the materials listed on Table 18.4 having a natural biological population, it still may be advisable in some cases for this

population to be supplemented by "designed" or pre-acclimated microorganisms to result in less start-up time and potentially more stable long term operational effectiveness.

5. CASE STUDIES

5.1. High Concentration 2-Propanol (IPA) and Acetone

It is often possible to continuously extend the range of biofiltration by use of high pre-acclimated microorganisms and extremophiles. For example to treat 25,000 m^3 of IPA and its intermediate acetone with 95% removal then, the design of such a biofilter is as follows:

A bench scale investigation reveals that it is possible to treat this stream with an inlet concentration of 15 g m^{-3} of IPA (2-propanol) with a final maximum EC of 280 g m^{-3} h^{-1}.

Thus, $C_I = 15\,g\,m^{-3}$ and so $C_O = 0.05 \times 15 = 0.75\,g\,m^{-3}$

Now,
$$EC = \frac{(C_I - C_O)\,A_f}{V_f} = 280 \tag{8}$$

And so
$$V = \frac{(C_I - C_O)A_f}{EC} = \frac{(15 - 0.75)25000}{280} = 1272\,m^3 \tag{9}$$

If we make the bed a typical depth of 1.5 m per stage and stack the bed two stages deep, then the cross-sectional area to treat this pollutant flow is:

$$A_{bed} = \frac{1272}{2 \times 1.5} = 424\,m^2 \tag{10}$$

and, in a square configuration this leads to a ~21 m × 21 m square bed.

For this type of operation, at a high EC and pollutant load a pre-acclimated microbial consortium would be needed (from the bench study), and an inert microbial support may therefore be an option. In this case the amount of support medium can be calculated as follows, based on a 0.45 voidage:

$$\text{Volume of packing} = (1 - 0.45) \times 1272 = 700\,m^3 \text{ of packing.}$$

A decision on mode of operating, such as up or downflow of the polluted air and method of delivery of liquid/nutrients, would subsequently need to be decided.

5.2. General Odor Control at a Municipal Wastewater Treatment Facility

The following is a case study of a successful biofilter application for odor control of a low concentration, but chemically diverse airstream. The case study describes the reasons for an air treatment system, the cost comparisons for a competing technology, a description of the decision-making process, and outcomes of the process. It should be noted that there are numerous ways to go about choosing your treatment system and this is one of many possible successful routes. However, this example does illustrate the great potential cost savings of biofiltration technology. Additional information is presented in Tables 18.7–18.11 and Figure 18.5.

Table 18.7
Microbial, pH, and moisture content averages during the pilot study

	Microbial Count	pH of Solid Support	Solid Support Moisture Content
July Average	1.1 E8 CFU/g	6.2	35% (w/w)
August Average	2.7 E5 CFU/g	6.6	30%
September Average	2.9 E6 CFU/g	6.7	28%
Take Down	3.4 E6 CFU/g	7.1	31%

Table 18.8
Air-borne chemicals monitored during the pilot study

Parameter	Overall Removal	Influent Mean (ppm)	Effluent Mean (ppm)
VOCs	93.7%	16.33 (±5.39)	0.94 (±5.72)
H_2S	100%[a]	0.06 (±0.22)	0.00 (±0.00)
Ammonia	81.6%	19.67 (±6.50)	3.61 (±5.14)

[a] Very low concentrations of H_2S.

Table 18.9
The estimated and actual costs of the two air treatment systems

	Chemical System	Biofilter
Estimated Capital Cost	$1,224,000	$941,000
Estimated Annual O&M	$194,000	$45,000
Actual Capital Cost	n/a	$1,120,000[b]
Actual O&M (1st year)[a]	n/a	$45,000

[a] Unit has been operating for 1.5 years.
[b] Includes all engineering, lava rock, etc.

The Neenah-Menasha Sewerage Commission owns and operates a regional 13 MGD wastewater treatment facility serving a population of 55,000 in Northeast Wisconsin. The plant serves the cities of Neenah and Menasha, Waverly Sanitary District, Town of Menasha Utility District, and Town of Neenah Sanitary District. Major industrial contributors to the plant include U.S. Papers, Gilbert Papers, Galloway Dairy and 20 pretreatment regulated industries. The treatment facility is located on the shore of Little Lake Butte des Morts and is surrounded on the remaining three sides by residential homes and a city park. Approximately 100 homes are within a 500-foot radius of the facility. The facility was originally constructed in the 1930s. Shortly after start-up of an expanded facility in 1986, residents began complaining about odors. In 1990, the commission authorized an odor survey that determined the main source of odors to be from the headworks and biosolids dewatering area.

Two types of vapor phase odor control technologies were given serious consideration: wet chemical scrubbing and biofiltration. Wet chemical scrubbing is very effective in removing ammonia, hydrogen sulfide and organic related odors. However, the major challenge in the design of a wet chemical scrubber is minimization of chemical use and cost. Multistage

Table 18.10
Parameters Monitored During Full-scale Operation

Parameter	Overall Removal	Influent Mean (ppm)	Effluent Mean (ppm)	Overall Average
VOCs	65%	32	11	n/a
H_2S	100%[a]	0.5	0	n/a
NH_3	100%	14	0	n/a
Solid Support Microorganisms	n/a	n/a	n/a	3.1 E6 CFU/g
Solid Support Moisture Content	n/a	n/a	n/a	29%[b]
pH of Solid Support	n/a	n/a	n/a	6.9
Air Flow Rate	n/a	n/a	n/a	46,500 cfm
Influent Air Relative Humidity	n/a	n/a	n/a	99+%
Influent Air Temperature	n/a	n/a	n/a	52–85°F[c]
TBRT	n/a	n/a	n/a	40 seconds

[a] Very low concentrations of H_2S.
[b] 90% of water holding capacity of the solid support.
[c] Temperature range. Largely dependent on season.

systems accomplish this best but these systems are still slaves to stoichiometry. Although effective, operation and maintenance (O&M) costs are high as well as capital costs. Biofiltration was also considered. Since our objective was a reliable low O&M cost system, biofiltration was a viable alternative.

Two vendor-offered biofilter systems, each with proprietary media and guaranteed performance, were considered. Vendor "A" offered a combination of a proprietary mixture of organic material (estimated three- to five-year life) installed over ceramic balls. The estimated capital cost was $675,000 to $1,125,000 plus engineering, installation and ducting. Media replacement cost was $75/cu. yd. Vendor "B" offered a specially engineered compost media with a five-year guarantee and 12–20% the overall size as a typical biofilter. Their media replacement cost was $200/cu. yd. Neither option was desirable to the client.

Rather than proceed with a "turnkey" vendor-supplied biofilter system, the client chose to characterize the odor constituents in the airstream and to pilot test a biofilter to demonstrate the system's performance. Lava rock was selected as the media because of its potential for long life, thus significantly reducing O&M expenditures.

After over four months of operation, the 56 ft^3 pilot-scale biofilter showed excellent performance. There was no visible degradation of the solid support and biomass levels were consistent, which indicated that lava rock would likely be an effective long-term solid support. While there was data collected on VOCs, H_2S, and NH_3 there was also a more subjective "smell test" performed by local residents, commissioners, and other interested parties. All of these tests demonstrated that the biofilter was effective in eliminating a significant portion of the objectionable odors in the airstream. While the ammonia, hydrogen sulfide, and VOCs were chosen for monitoring, it was impossible to determine what portion of the total odor these chemicals actually contribute. Since extensive air analysis work indicated there were

Table 18.11
Parameters that are monitored during various biofiltration applications

Parameter	Relative Importance[a]	Relative Cost[a]	Critical Information Provided
Concentrations/Removal of Target Compounds in the Airstream	3	4 or 5	Critical information to assess the 'performance' of the system. However, for complex odor applications complex monitoring may not be as valuable as the smell test at the site. This is more critical for applications where total emissions are part of a permitting or discharge process. Often the most costly and requires the most capital equipment of the monitoring parameters.
Microbial Counts	2	3	Since the biofilter is a living system, this is often a cost-effective method to assess the overall health of the system. Large increases in numbers can be problematic as it may result in clogging of the system. Large decreases in counts may indicate an accumulation of toxic intermediates, changes in the airstream, or a lack of nutrients or moisture. Counts are generally greater than 1.0×10^6 Colony Forming Units (CFU) per gram of solid support.
Moisture Content of the Solid Support	1	2	Inexpensive parameter to monitor and critical to good chemical partitioning and microbial growth. It is usually desirable to have the moisture content stable and as close to the moisture-holding capacity of the solid support as possible.
Nutrients (N and P)	1	3	Critical to the proper growth of microorganisms in the biofilter. Proper nutrient levels have been shown to be a critical factor in efficient biofilter operation. Usually samples are collected and sent to a laboratory that does these analyses, thus making it a relatively easy parameter to monitor.
pH of Solid Support	1	2	pH of the solid support is very important to monitor due to potential acidic intermediates which are produced by biological oxidations. Most biofilters operate in a pH range of 6–8.

(*Continued*)

Table 18.11 (*Continued*)

Parameter	Relative Importance[a]	Relative Cost[a]	Critical Information Provided
Porosity/Integrity of the Solid Support	4 or 1[b]	2	The porosity of the solid support is important to determine to calculate the actual residence time in the system. It is critical to monitor this parameter for natural solid support materials as they will degrade over time. By catching the degradation of this material early it may help the operator avoid system failures and unexpected down time.
Pressure Drop Across Biofilter bed	2	1	One of the most critical factors to monitor and very inexpensive. Increases in pressure drop across a biofilter bed can indicate microbial overgrowth on the solid support. This overgrowth can lead to log-order increases in microbial growth and increasing pressure drops. These large pressure drops can lead to large increases in electrical costs due to increased work by the motors to move air through the system. Since these electrical costs are often one of the largest operating expenses, large pressure drops can drastically increase operational costs. Large pressure drops can be a prelude to complete system failure.
Relative Humidity (RH) of the Influent Air	2	1	This is an inexpensive parameter to monitor. Since the influent portion of many large-scale systems is difficult to access, this assures the operator the influent zone possesses adequate moisture. Influent air should be 99.9% RH for best operation.
Temperature	1	1	Easy and inexpensive parameter to monitor. Used to help assess if temperature changes can be a contributing factor to changes in biofilter performance. Generally, the closer to 70°F the influent air is, the better performance your system will have.

[a] Relative scale is 1–5 with 5 being the most important or costly and 1 being the least important or costly.
[b] This is very important (4) if the solid support is a natural material like wood chips or bark. However, it is less important for materials that do not breakdown readily like many of the synthetic solid supports.

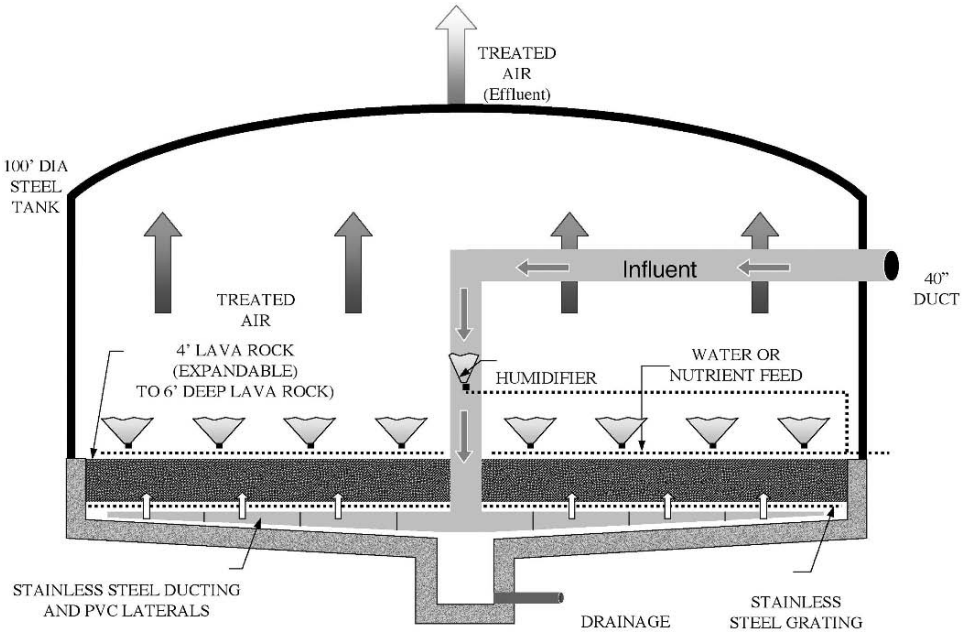

Fig. 18.5. Full-Scale Biofilter design from the above Case Study (*Source:* G. Kleinheinz, unpublished).

hundreds of chemicals in the airstream, the client chose also to conduct subjective tests for odor removal. Since each chemical in the complex airstream has a different dispersion rate in air, odor threshold, it would be nearly impossible to characterize the removal of each of these chemicals from the pilot-scale system. Since each person defines "odor" differently, the client thought it was important to gain input on the pilot-scale system from local residents (who initially complained of the odors) and from the commissioners who will decide on funding for a full-scale system. All residents who smell-tested the system agreed it significantly reduced the odors from the airstream.

Due to the success of the pilot-scale biofilter, the cost of a 45,000 cfm chemical scrubber was compared to a lava rock-based biofilter. The biofilter was to be constructed in two existing unused 100-ft diameter steel tanks with an existing concrete floor/foundation and aluminum cover. The estimated and actual costs are compared below:

Based upon these costs and the pilot-scale demonstration, the biofilter system was selected. Figure 18.6 shows a cutaway view of the basic design. Each existing steel tank was retrofitted to hold 4-feet-deep lava rock media. Stainless steel grating was used to support the rock. PVC piping was used to distribute the foul air throughout the tank floor. A spray system using non-potable water was used to keep the lava rock moist. Approximately 20 gpm/unit was provided to keep the lava rock moist. All drainage was collected and returned to the headworks for treatment. A chemical feed pump was provided to allow for the addition of nutrients if needed. The biofilter exhaust would exit the biofilter through the hatch openings on the aluminum covers. At a flow rate of $\sim$45,000 cfm, the units were sized to have an approximate empty

bed residence time of 1.4 minutes. The lava rock has a porosity of approximately 50% for an actual residence time of approximately 42 seconds. Based on pilot testing, this should allow for further air handling capacity in the future if needed.

The pilot test results demonstrated that biofiltration was a capital cost competitive, low O&M cost solution for effective odor control at this site. The biofilter provided the added benefit (over chemical scrubbing) of not requiring the on-site storage of large amounts of toxic chemicals. Test data allowed for a properly sized system specific to the odor constituents, rather than force-fitting a vendor system to the site. The pilot test allowed for local politicians and area residents to sample the air quality from the unit, which allowed their buy-in to the technology.

The unit has operated for over two years with virtually no odor complaints from local residents. Given the relatively low cost of operation and the success in terms of public relations and odor mitigation, this application of biofiltration has been a success.

6. PROCESS CONTROL AND MONITORING

As these systems contain living entities, it is vital that proper process monitoring is carried out to ensure the long-term stability of the process. For example, if the bed dries out too much, and/or high concentrations of pollutant or extremes of temperature or pH are experienced, then this may lead to a severe decrease in performance, or in the worst case complete bed failure (i.e. pollutant breakthrough). The control and monitoring of biofiltration systems is highly variable, from little if any monitoring to complete monitoring of all operational and process parameters. Since the process at the facility usually dictates the airflow rate, it is often not considered a controllable variable. However, it is often important to monitor flow rates to verify that fans and the distribution system are operating properly. Generally, the more monitoring conducted, the more the operator understands about the treatment system. More importantly, the more monitoring that is conducted, the more likely that the operator will identify any upsets or changes in the system before they become operations problems that can lead to downtime. By identifying potential issues early, it is easier to correct them prior to serious damage to the microbes or equipment in the system. While extensive monitoring is a "best case" scenario, it is often not practical or economical for some facilities. In these cases, the operator must make changes regarding which parameters to.

Often the cost-to-need ratio dictates the level of monitoring that is performed at a site. That is, if the biofiltration is for odor control only and the facility has a relatively small air treatment budget, it may choose to do minimal monitoring. Conversely, if a biofilter is being used to treat chemicals that are a regulated discharge such as some VOCs, it may be more important for the facility to monitor the system more stringently. When a system used for odor control goes out of service, it often leads to some odor complaints for the operator, but few regulatory problems. When a system treating regulated chemicals goes out of service, it may mean the facility will exceed its discharge permit and this could cause the facility to shut down or pay fines to exceed permit discharge levels.

Table 18.11 shows some parameters that are often monitored and the information that the monitoring provides the operator. If there is a need to compromise on some monitoring,

the operator should use the information that is known about the process stream to help decide which parameters are most critical for that application. For example, if your system were treating a significant amount of reduced sulfur compounds, pH would be a critical factor to monitor due to the large amount of toxic products produced by the oxidation of reduced sulfur compounds. In general, for biofilter systems, moisture and pH distributions are vital pieces of information.

7. LIMITATIONS OF THE TECHNOLOGY

As discussed previously there are several "traditional" limitations to biofiltration technology such as high concentrations of chemicals, size of some units, microbial capabilities, and process air temperatures. However, while these have been traditional limitations, recent work in both biofilter design and operation has helped overcome some of these problems of the past.

As mentioned before, operators are largely responsible (along with designers) for successful operation of biofilters. A biofilter operator needs to be aware of the operation parameters of the system to avoid such issues as drying out of the bed, compaction and overgrowth of microorganisms, and pH decrease to name just a few.

As more is being understood about microbial population dynamics in these systems, the operational window for biofiltration systems is continually widening. It is imperative that, when biofiltration is being considered as a treatment technology, all factors are considered prior to design and start-up so some potential limitations can be overcome. By using knowledgeable planning, the success stories of biofiltration will continue to expand in both total number and diversity of applications.

8. CONCLUSIONS

Biofiltration technologies are gaining wider acceptance as a viable air treatment technology. Biofilters are not applicable to all airstreams, however recent development of biofiltration technology has seen an ever-increasing range of applications. Recent research and development of these systems has led to a better understanding of sizing, operational, and microbiological aspects of the treatment process. Biofilters are no longer the "black box" in which treatment takes place. We are now able to understand the complex chemical and biological interaction that takes place in these systems to better design them for a myriad of applications which were previously not considered appropriate applications of biofiltration.

It is imperative that biofilters be sized and properly fitted to their intended application. Too often, one biofilter design is adapted to many different applications with less than satisfactory results. While the same design may be applicable for several applications it is important that each application is evaluated on its needs and specific characteristics. These characteristics include airflow to be treated, concentration of chemicals in the airstream, temperature of the airstream, biodegradability of the contaminants, etc. Once these considerations, and possibly others, are evaluated, the choice to go with a biofilter then can be made. Once biofilters are decided upon as the treatment method, the designers can work on sizing, geometry, solid support material, etc. depending on the characteristics of the airstream. It is imperative that

the unit be properly installed and "fit" to the specific application. A bench or pilot scale trial is highly recommended in this context.

Once the biofilter is operational, a monitoring protocol must be implemented which allows for the evaluation of performance and for the notification of the operator of any upsets in the system. Since these are biological systems, it is imperative to find small problems before they become large problems that require downtime of the system.

In principle, biofilters are very simple methods of air treatment. However, increased understanding of the engineering and microbiology involved in the process has made them one of the more difficult treatments systems to operate effectively. That is, it takes a good understanding of engineering, the process stream being treated, and the microbiology in the system to allow for the long-term operation of these systems. If properly designed, operated, monitored, and maintained, a biofilter should allow for many years of cost-effective air treatment. This cost-effective operation will likely save the operator a significant amount (tens to hundreds of thousands of dollars, or more) in operational costs over its lifetime when compared to alternative treatment technologies (27–30).

NOMENCLATURE

A = Area (ft^2 or m^2)
C_i = Influent concentration (g/m^3 or lb/ft^3)
C_o = Effluent concentration (g/m^3 or lb/ft^3)
V_b = Volume of biofilter bed (m^3 or ft^3)
A_r = Airflow rate (m^3/minute or cfm)
$EBRT$ = Empty bed residence time (seconds or minutes)
$TBRT$ = True bed residence time (seconds or minutes)
V_{ss} = Volume of solid support (m^3 or ft^3)
M_p = Media porosity (%)
V_s = Volume of a given space (m^3 or ft^3)
VL = Volumetric Loading
RE = Removal Efficiency
EC = Elimination Capacity
Δp_{bed} = Pressure drop across bed (kPa or psi)

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Appendix: Conversion Factors for Environmental Engineers

Lawrence K. Wang

CONTENTS

Abstract With the current trend toward metrication, the question of using a consistent system of units has been a problem. Wherever possible, the authors of this *Handbook of Environmental Engineering* series have used the British system (fps) along with the metric equivalent (mks, cgs, or SIU) or vice versa. For the convenience of the readers around the world, this book provides a detailed Conversion Factors for Environmental Engineers. In addition, the basic and supplementary units, the derived units and quantities, important physical constants, the properties of water, and the Periodic Table of the elements, are also presented in this document.

Key Words Conversion factors • British units • metric units • physical constants • water properties • periodic table of the elements • environmental engineers • Lenox Institute of Water Technology • mks (meter-kilogram-second) • cgs (centimeter-gram-second) • SIU (Système international d'unités; International System of Units) • fps (foot-pound-second).

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Edited by: L. K. Wang et al. © The Humana Press, Totowa, NJ

1. CONSTANTS AND CONVERSION FACTORS

Multiply	*by*	*to obtain*
abamperes	10	amperes
abamperes	2.99796×10^{10}	statamperes
abampere-turns	12.566	gilberts
abcoulombs	10	coulombs (abs)
abcoulombs	2.99796×10^{10}	statcoulombs
abcoulombs/kg	30,577	statcoulombs/dyne
abfarads	1×10^{9}	farads (abs)
abfarads	8.98776×10^{20}	statfarads
abhenries	1×10^{-9}	henries (abs)
abhenries	1.11263×10^{-21}	stathenries
abohms	1×10^{-9}	ohms (abs)
abohms	1.11263×10^{-21}	statohms
abvolts	3.33560×10^{-11}	statvolts
abvolts	1×10^{-8}	volts (abs)
abvolts/centimeters	2.540005×10^{-8}	volts (abs)/inch
acres	0.4046	ha
acres	43,560	square feet
acres	4047	square meters
acres	1.562×10^{-3}	square miles
acres	4840	square yards
acre-feet	43,560	cubic feet
acre-feet	1233.5	cubic meters
acre-feet	325,850	gallons (U.S.)
amperes (abs)	0.1	abamperes
amperes (abs)	1.036×10^{-5}	faradays/second
amperes (abs)	2.9980×10^{9}	statamperes
ampere-hours (abs)	3600	coulombs (abs)
ampere-hours	0.03731	faradays
amperes/sq cm	6.452	amps/sq in
amperes/sq cm	10^{4}	amps/sq meter
amperes/sq in	0.1550	amps/sq cm
amperes/sq in	1550.0	amps/sq meter
amperes/sq meter	10^{-4}	amps/sq cm
amperes/sq meter	6.452×10^{-4}	amps/sq in
ampere-turns	1.257	gilberts
ampere-turns/cm	2.540	amp-turns/in
ampere-turns/cm	100.0	amp-turns/meter
ampere-turns/cm	1.257	gilberts/cm
ampere-turns/in	0.3937	amp-turns/cm
ampere-turns/in	39.37	amp-turns/meter
ampere-turns/in	0.4950	gilberts/cm

Multiply	*by*	*to obtain*
ampere-turns/meter	0.01	amp-turns/cm
ampere-turns/meter	0.0254	amp-turns/in
ampere-turns/meter	0.01257	gilberts/cm
angstrom units	1×10^{-8}	centimeters
angstrom units	3.937×10^{-9}	inches
angstrom unit	1×10^{-10}	meter
angstrom unit	1×10^{-4}	micron or μm
ares	0.02471	acre (U.S.)
ares	1076	square feet
ares	100	square meters
ares	119.60	sq yards
assay tons	29.17	grams
astronomical unit	1.495×10^{8}	kilometers
atmospheres (atm)	0.007348	tons/sq inch
atmospheres	76.0	cms of mercury
atmospheres	1.01325×10^{6}	dynes/square centimeter
atmospheres	33.90	ft of water (at 4°C)
atmospheres	29.92	inches of mercury (at 0°C)
atmospheres	1.033228	kg/sq cm
atmospheres	10,332	kg/sq meter
atmospheres	760.0	millimeters of mercury
atmospheres	14.696	pounds/square inch
atmospheres	1.058	tons/sq foot
avograms	1.66036×10^{-24}	grams
bags, cement	94	pounds of cement
barleycorns (British)	1/3	inches
barleycorns (British)	8.467×10^{-3}	meters
barrels (British, dry)	5.780	cubic feet
barrels (British, dry)	0.1637	cubic meters
barrels (British, dry)	36	gallons (British)
barrels, cement	170.6	kilograms
barrels, cement	376	pounds of cement
barrels, cranberry	3.371	cubic feet
barrels, cranberry	0.09547	cubic meters
barrels, oil	5.615	cubic feet
barrels, oil	0.1590	cubic meters
barrels, oil	42	gallons (U.S.)
barrels, (U.S., dry)	4.083	cubic feet
barrels (U.S., dry)	7056	cubic inches
barrels (U.S., dry)	0.11562	cubic meters
barrels (U.S., dry)	105.0	quarts (dry)
barrels (U.S., liquid)	4.211	cubic feet
barrels (U.S., liquid)	0.1192	cubic meters
barrels (U.S., liquid)	31.5	gallons (U.S.)

Multiply	*by*	*to obtain*
bars	0.98692	atmospheres
bars	10^6	dynes/sq cm
bars	1.0197×10^4	kg/sq meter
bars	1000	millibar
bars	750.06	mm of Hg (0°C)
bars	2089	pounds/sq ft
bars	14.504	pounds/sq in
barye	1.000	dynes/sq cm
board feet	1/12	cubic feet
board feet	144 sq.in. × 1 in.	cubic inches
boiler horsepower	33,475	BTU (mean)/hour
boiler horsepower	34.5	pounds of water evaporated from and at 212°F (per hour)
bolts (U.S., cloth)	120	linear feet
bolts (U.S., cloth)	36.576	meters
bougie decimales	1	candles (int)
BTU (mean)	251.98	calories, gram (g. cal)
BTU (mean)	0.55556	centigrade heat units (chu)
BTU (mean)	1.0548×10^{10}	ergs
BTU (mean)	777.98	foot-pounds
BTU (mean)	3.931×10^{-4}	horsepower-hrs (hp-hr)
BTU (mean)	1055	joules (abs)
BTU (mean)	0.25198	kilograms, cal (kg cal)
BTU (mean)	107.565	kilogram-meters
BTU (mean)	2.928×10^{-4}	kilowatt-hr (Kwh)
BTU (mean)	10.409	liter-atm
BTU (mean)	6.876×10^{-5}	pounds of carbon to CO_2
BTU (mean)	0.29305	watt-hours
BTU (mean)/cu ft	37.30	joule/liter
BTU/hour	0.2162	foot-pound/sec
BTU/hour	0.0700	gram-cal/sec
BTU/hour	3.929×10^{-4}	horsepower-hours (hp-hr)
BTU/hour	0.2930711	watt (w)
BTU/hour (feet)°F	1.730735	joule/sec (m)°k
BTU/hour (feet^2)	3.15459	joule/m^2-sec
BTU (mean)/hour(feet^2)°F	1.3562×10^{-4}	gram-calorie/second (cm^2)°C
BTU (mean)/hour(feet^2)°F	3.94×10^{-4}	horsepower/(ft^2)°F
BTU (mean)/hour(feet^2)°F	5.678264	joule/sec (m^2)°k
BTU (mean)/hour(feet^2)°F	4.882	kilogram-calorie/hr (m^2)°C
BTU (mean)/hour(feet^2)°F	5.682×10^{-4}	watts/(cm^2)°C
BTU (mean)/hour(feet^2)°F	2.035×10^{-3}	watts/(in^2)°C
BTU (mean)/(hour)(feet^2) (°F/inch)	3.4448×10^{-4}	calories, gram (15°C)/sec (cm^2) (°C/cm)
BTU (mean)/(hour)(feet^2) (°F/in.)	1	chu/(hr)(ft^2)(°C/in)

Multiply	*by*	*to obtain*
BTU (mean)/(hour)(feet^2) (°F/inch)	1.442×10^{-3}	joules (abs)/(sec)(cm^2) (°C/cm)
BTU (mean)/(hour)(feet^2) (°F/inch)	1.442×10^{-3}	watts/(cm^2) (°C/cm)
BTU/min	12.96	ft lb/sec
BTU/min	0.02356	hp
BTU/min	0.01757	kw
BTU/min	17.57	watts
BTU/min/ft^2	0.1221	watts/sq inch
BTU/pound	0.5556	calories-gram(mean)/gram
BTU/pound	0.555	kg-cal/kg
BTU/pound/°F	1	calories, gram/gram/°C
BTU/pound/°F	4186.8	joule/kg/°k
BTU/second	1054.350	watt (W)
buckets (British, dry)	1.818×10^4	cubic cm
buckets (British, dry)	4	gallons (British)
bushels (British)	1.03205	bushels (U.S.)
bushels (British)	1.2843	cubic feet
bushels (British)	0.03637	cubic meters
bushels (U.S.)	1.2444	cubic feet
bushels (U.S.)	2150.4	cubic inch
bushels (U.S.)	0.035239	cubic meters
bushels (U.S.)	35.24	liters (L)
bushels (U.S.)	4	pecks (U.S.)
bushels (U.S.)	64	pints (dry)
bushels (U.S.)	32	quarts (dry)
butts (British)	20.2285	cubic feet
butts (British)	126	gallons (British)
cable lengths	720	feet
cable lengths	219.46	meters
calories (thermochemical)	0.999346	calories (Int. Steam Tables)
calories, gram (g. cal or simply cal.)	3.9685×10^{-3}	BTU (mean)
calories, gram (mean)	0.001459	cubic feet atmospheres
calories, gram (mean)	4.186×10^7	ergs
calories, gram (mean)	3.0874	foot-pounds
calories, gram (mean)	4.186	joules (abs)
calories, gram (mean)	0.001	kg cal (calories, kilogram)
calories, gram (mean)	0.42685	kilograms-meters
calories, gram (mean)	0.0011628	watt-hours
calories, gram (mean)/gram	1.8	BTU (mean)/pound
cal/gram-°C	4186.8	joule/kg-°k
candle power (spherical)	12.566	lumens
candles (int)	0.104	carcel units
candles (int)	1.11	hefner units
candles (int)	1	lumens (int)/steradian
candles (int)/square centimeter	2919	foot-lamberts

Multiply	*by*	*to obtain*
candles (int)/square centimeter	3.1416	lamberts
candles (int)/square foot	3.1416	foot-lamberts
candles (int)/square foot	3.382×10^{-3}	lamberts
candles (int)/square inch	452.4	foot-lamberts
candles (int)/square inch	0.4870	lamberts
candles (int)/square inch	0.155	stilb
carats (metric)	3.0865	grains
carats (metric)	0.2	grams
centals	100	pounds
centares (centiares)	1.0	sq meters
centigrade heat units (chu)	1.8	BTU
centigrade heat units (chu)	453.6	calories, gram (15°C)
centigrade heat units (chu)	1897.8	joules (abs)
centigrams	0.01	grams
centiliters	0.01	liters
centimeters	0.0328083	feet (U.S.)
centimeters	0.3937	inches (U.S.)
centimeters	0.01	meters
centimeters	6.214×10^{-6}	miles
centimeters	10	millimeters
centimeters	393.7	mils
centimeters	0.01094	yards
cm of mercury	0.01316	atm
cm of mercury	0.4461	ft of water
cm of mercury	136.0	kg/square meter
cm of mercury	1333.22	newton/meter^2 (N/m^2)
cm of mercury	27.85	psf
cm of mercury	0.1934	psi
cm of water (4°C)	98.0638	newton/meter^2 (N/m^2)
centimeters-dynes	1.020×10^{-3}	centimeter-grams
centimeter-dynes	1.020×10^{-8}	meter-kilograms
centimeter-dynes	7.376×10^{-8}	pound-feet
centimeter-grams	980.7	centimeter-dynes
centimeter-grams	10^{-5}	meter-kilograms
centimeter-grams	7.233×10^{-5}	pound-feet
centimeters/second	1.969	fpm (ft/min)
centimeters/second	0.0328	fps (ft/sec)
centimeters/second	0.036	kilometers/hour
centimeters/second	0.1943	knots
centimeters/second	0.6	m/min
centimeters/second	0.02237	miles/hour
centimeters/second	3.728×10^{-4}	miles/minute
cms/sec./sec.	0.03281	feet/sec/sec
cms/sec./sec.	0.036	kms/hour/sec

Multiply	*by*	*to obtain*
cms/sec./sec.	0.02237	miles/hour/sec
centipoises	3.60	kilograms/meter hour
centipoises	10^{-3}	kilograms/meter second
centipoises	0.001	newton-sec/m^2
centipoises	2.089×10^{-5}	pound force second/square foot
centipoises	2.42	pounds/foot hour
centipoises	6.72×10^{-4}	pounds/foot second
centistoke	1.0×10^{-6}	$meter^2$/sec
chains (engineers' or Ramden's)	100	feet
chains (engineers' or Ramden's)	30.48	meters
chains (surveyors' or Gunter's)	66	feet
chains (surveyors' or Gunter's)	20.12	meters
chaldrons (British)	32	bushels (British)
chaldrons (U.S.)	36	bushels (U.S.)
cheval-vapours	0.9863	horsepower
cheval-vapours	735.5	watts (abs)
cheval-vapours heures	2.648×10^6	joules (abs)
chu/(hr)(ft^2)(°C/in.)	1	BTU/(hr)(ft^2)(°F/in.)
circular inches	0.7854	square inches
circular millimeters	7.854×10^{-7}	square meters
circular mils	5.067×10^{-6}	square centimeters
circular mils	7.854×10^{-7}	square inches
circular mils	0.7854	square mils
circumferences	360	degrees
circumferences	400	grades
circumferences	6.283	radians
cloves	8	pounds
coombs (British)	4	bushels (British)
cords	8	cord feet
cords	$8' \times 4' \times 4'$	cubic feet
cords	128	cubic feet
cords	3.625	cubic meters
cord-feet	$4' \times 4' \times 1'$	cubic feet
coulombs (abs)	0.1	abcoulombs
coulombs (abs)	6.281×10^{18}	electronic charges
coulombs (abs)	2.998×10^9	statcoulombs
coulombs (abs)	1.036×10^{-5}	faradays
coulombs/sq cm	64.52	coulombs/sq in
coulombs/sq cm	10^4	coulombs/sq meter
coulombs/sq in	0.1550	coulombs/sq cm
coulombs/sq in	1550	coulombs/sq meter
coulombs/sq meter	10^{-4}	coulombs/sq cm
coulombs/sq meter	6.452×10^{-4}	coulombs/sq in
cubic centimeters	3.531445×10^{-5}	cubic feet (U.S.)

Multiply	*by*	*to obtain*
cubic centimeters	6.102×10^{-2}	cubic inches
cubic centimeters	10^{-6}	cubic meters
cubic centimeters	1.308×10^{-6}	cubic yards
cubic centimeters	2.6417×10^{-4}	gallons (U.S.)
cubic centimeters	0.001	liters
cubic centimeters	0.033814	ounces (U.S., fluid)
cubic centimeters	2.113×10^{-3}	pints (liq.)
cubic centimeters	1.057×10^{-3}	quarts (liq.)
cubic feet (British)	0.9999916	cubic feet (U.S.)
cubic feet (U.S.)	0.8036	bushels (dry)
cubic feet (U.S.)	28317.016	cubic centimeters
cubic feet (U.S.)	1728	cubic inches
cubic feet (U.S.)	0.02832	cubic meters
cubic feet (U.S.)	0.0370	cubic yard
cubic feet (U.S.)	7.48052	gallons (U.S.)
cubic feet (U.S.)	28.31625	liters
cubic feet (U.S.)	59.84	pints (liq.)
cubic feet (U.S.)	29.92	quarts (liq.)
cubic feet of common brick	120	pounds
cubic feet of water (60°F)	62.37	pounds
cubic foot-atmospheres	2.7203	BTU (mean)
cubic foot-atmospheres	680.74	calories, gram (mean)
cubic foot-atmospheres	2116	foot-pounds
cubic foot-atmospheres	2869	joules (abs)
cubic foot-atmospheres	292.6	kilogram-meters
cubic foot-atmospheres	7.968×10^{-4}	kilowatt-hours
cubic feet/hr	0.02832	m^3/hr
cubic feet/minute	472.0	cubic cm/sec
cubic feet/minute	1.6992	cu m/hr
cubic feet/minute	0.0283	cu m/min
cubic feet/minute	0.1247	gallons/sec
cubic feet/minute	0.472	liter/sec
cubic feet/minute	62.4	lbs of water/min
cubic feet/min/1000 cu ft	0.01667	liter/sec/cu m
cubic feet/second	1.9834	acre-feet/day
cubic feet/second	1.7	cu m/min
cubic feet/second	0.02832	m^3/sec
cubic feet/second	448.83	gallons/minute
cubic feet/second	1699	liter/min
cubic feet/second	28.32	liters/sec
cubic feet/second (cfs)	0.64632	million gallons/day (MGD)
cfs/acre	0.07	m^3/sec-ha
cfs/acre	4.2	cu m/min/ha
cfs/sq mile	0.657	cu m/min/sq km

Multiply	*by*	*to obtain*
cubic inches (U.S.)	16.387162	cubic centimeters
cubic inches (U.S.)	5.787×10^{-4}	cubic feet
cubic inches (U.S.)	1.0000084	cubic inches (British)
cubic inches (U.S.)	1.639×10^{-5}	cubic meters
cubic inches (U.S.)	2.143×10^{-5}	cubic yards
cubic inches (U.S.)	4.329×10^{-3}	gallons (U.S.)
cubic inches (U.S.)	1.639×10^{-2}	liters
cubic inches (U.S.)	16.39	mL
cubic inches (U.S.)	0.55411	ounces (U.S., fluid)
cubic inches (U.S.)	0.03463	pints (liq.)
cubic inches (U.S.)	0.01732	quarts (liq.)
cubic meters	8.1074×10^{-4}	acre-feet
cubic meters	8.387	barrels (U.S., liquid)
cubic meters	28.38	bushels (dry)
cubic meters	10^6	cubic centimeters
cubic meters	35.314	cubic feet (U.S.)
cubic meters	61,023	cubic inches (U.S.)
cubic meters	1.308	cubic yards (U.S.)
cubic meters	264.17	gallons (U.S.)
cubic meters	1000	liters
cubic meters	2113	pints (liq.)
cubic meters (m^3)	1057	quarts (liq.)
cubic meters/day	0.183	gallons/min
cubic meters/ha	106.9	gallons/acre
cubic meters/hour	0.2272	gallons/minute
cubic meters/meter-day	80.53	gpd/ft
cubic meters/minute	35.314	cubic ft/minute
cubic meters/second	35.314	cubic ft/sec
cubic meters/second	22.82	MGD
cubic meters/sec-ha	14.29	cu ft/sec-acre
cubic meters/$meters^2$-day	24.54	gpd/ft^2
cubic yards (British)	0.9999916	cubic yards (U.S.)
cubic yards (British)	0.76455	cubic meters
cubic yards (U.S.)	7.646×10^5	cubic centimeters
cubic yards (U.S.)	27	cubic feet (U.S.)
cubic yards (U.S.)	46,656	cubic inches
cubic yards (U.S.)	0.76456	cubic meters
cubic yards (U.S.)	202.0	gallons (U.S.)
cubic yards (U.S.)	764.6	liters
cubic yards (U.S.)	1616	pints (liq.)
cubic yards (U.S.)	807.9	quarts (liq.)
cubic yards of sand	2700	pounds
cubic yards/minute	0.45	cubic feet/second
cubic yards/minute	3.367	gallons/second

Multiply	*by*	*to obtain*
cubic yards/minute	12.74	liters/second
cubits	45.720	centimeters
cubits	1.5	feet
dalton	1.65×10^{-24}	gram
days	1440	minutes
days	86,400	seconds
days (sidereal)	86164	seconds (mean solar)
debye units (dipole moment)	10^{18}	electrostatic units
decigrams	0.1	grams
deciliters	0.1	liters
decimeters	0.1	meters
degrees (angle)	60	minutes
degrees (angle)	0.01111	quadrants
degrees (angle)	0.01745	radians
degrees (angle)	3600	seconds
degrees/second	0.01745	radians/seconds
degrees/second	0.1667	revolutions/min
degrees/second	0.002778	revoltuions/sec
degree Celsius	$°F = (°C \times 9/5) + 32$	Fahrenheit
degree Celsius	$°K = °C + 273.15$	Kelvin
degree Fahrenheit	$°C = (°F - 32) \times 5/9$	Celsius
degree Fahrenheit	$°K = (°F + 459.67)/1.8$	Kelvin
degree Rankine	$°K = °R/1.8$	Kelvin
dekagrams	10	grams
dekaliters	10	liters
dekameters	10	meters
drachms (British, fluid)	3.5516×10^{-6}	cubic meters
drachms (British, fluid)	0.125	ounces (British, fluid)
drams (apothecaries' or troy)	0.1371429	ounces (avoirdupois)
drams (apothecaries' or troy)	0.125	ounces (troy)
drams (U.S., fluid or apoth.)	3.6967	cubic cm
drams (avoirdupois)	1.771845	grams
drams (avoirdupois)	27.3437	grains
drams (avoirdupois)	0.0625	ounces
drams (avoirdupois)	0.00390625	pounds (avoirdupois)
drams (troy)	2.1943	drams (avoirdupois)
drams (troy)	60	grains
drams (troy)	3.8879351	grams
drams (troy)	0.125	ounces (troy)
drams (U.S., fluid)	3.6967×10^{-6}	cubic meters
drams (U.S., fluid)	0.125	ounces (fluid)
dynes	0.00101972	grams

Multiply	*by*	*to obtain*
dynes	10^{-7}	joules/cm
dynes	10^{-5}	joules/meter (newtons)
dynes	1.020×10^{-6}	kilograms
dynes	1×10^{-5}	newton (N)
dynes	7.233×10^{-5}	poundals
dynes	2.24809×10^{-6}	pounds
dyne-centimeters (torque)	7.3756×10^{-8}	pound-feet
dynes/centimeter	1	ergs/square centimeter
dynes/centimeter	0.01	ergs/square millimeter
dynes/square centimeter	9.8692×10^{-7}	atmospheres
dynes/square centimeter	10^{-6}	bars
dynes/square centimeter	2.953×10^{-5}	inch of mercury at 0°C
dynes/square centimeter	4.015×10^{-4}	inch of water at 4°C
dynes/square centimeter	0.01020	kilograms/square meter
dynes/square centimeter	0.1	newtons/square meter
dynes/square centimeter	1.450×10^{-5}	pounds/square inch
electromagnetic fps units of magnetic permeability	0.0010764	electromagnetic cgs units of magnetic permeability
electromagnetic fps units of magnetic permeability	1.03382×10^{-18}	electrostatic cgs units of magnetic permeability
electromagnetic cgs units, of magnetic permeability	1.1128×10^{-21}	electrostatic cgs units of magnetic permeability
electromagnetic cgs units of mass resistance	9.9948×10^{-6}	ohms (int)-meter-gram
electronic charges	1.5921×10^{-19}	coulombs (abs)
electron-volts	1.6020×10^{-12}	ergs
electron-volts	1.0737×10^{-9}	mass units
electron-volts	0.07386	rydberg units of energy
electronstatic cgs units of Hall effect	2.6962×10^{31}	electromagnetic cgs units of Hall effect
electrostatic fps units of charge	1.1952×10^{-6}	coulombs (abs)
electrostatic fps units of magnetic permeability	929.03	electrostatic cgs units of magnetic permeability
ells	114.30	centimeters
ells	45	inches
ems, pica (printing)	0.42333	centimeters
ems, pica (printing)	1/6	inches
ergs	9.4805×10^{-11}	BTU (mean)
ergs	2.3889×10^{-8}	calories, gram (mean)
ergs	1	dyne-centimeters
ergs	7.3756×10^{-8}	foot-pounds
ergs	0.2389×10^{-7}	gram-calories
ergs	1.020×10^{-3}	gram-centimeters

Multiply	*by*	*to obtain*
ergs	3.7250×10^{-14}	horsepower-hrs
ergs	10^{-7}	joules (abs)
ergs	2.390×10^{-11}	kilogram-calories (kg cal)
ergs	1.01972×10^{-8}	kilogram-meters
ergs	0.2778×10^{-13}	kilowatt-hrs
ergs	0.2778×10^{-10}	watt-hours
ergs/second	5.692×10^{-9}	BTU/min
ergs/second	4.426×10^{-6}	foot-pounds/min
ergs/second	7.376×10^{-8}	foot-pounds/sec
ergs/second	1.341×10^{-10}	horsepower
ergs/second	1.434×10^{-9}	kg-calories/min
ergs/second	10^{-10}	kilowatts
farad (international of 1948)	0.9995	farad (F)
faradays	26.80	ampere-hours
faradays	96,500	coulombs (abs)
faradays/second	96,500	amperes (abs)
farads (abs)	10^{-9}	abfarads
farads (abs)	10^{6}	microfarads
farads (abs)	8.9877×10^{11}	statfarads
fathoms	6	feet
fathom	1.829	meter
feet (U.S.)	1.0000028	feet (British)
feet (U.S.)	30.4801	centimeters
feet (U.S.)	12	inches
feet (U.S.)	3.048×10^{-4}	kilometers
feet (U.S.)	0.30480	meters
feet (U.S.)	1.645×10^{-4}	miles (naut.)
feet (U.S.)	1.893939×10^{-4}	miles (statute)
feet (U.S.)	304.8	millimeters
feet (U.S.)	1.2×10^{4}	mils
feet (U.S.)	1/3	yards
feet of air (1 atmosphere, 60°F)	5.30×10^{-4}	pounds/square inch
feet of water	0.02950	atm
feet of water	0.8826	inches of mercury
feet of water at 39.2°F	0.030479	kilograms/square centimeter
feet of water at 39.2°F	2988.98	newton/meter^2 (N/m^2)
feet of water at 39.2°F	304.79	kilograms/square meter
feet of water	62.43	pounds/square feet (psf)
feet of water at 39.2°F	0.43352	pounds/square inch (psi)
feet/hour	0.08467	mm/sec
feet/min	0.5080	cms/sec
feet/min	0.01667	feet/sec
feet/min	0.01829	km/hr
feet/min	0.3048	meters/min

Multiply	*by*	*to obtain*
feet/min	0.01136	miles/hr
feet/sec	30.48	cm/sec
feet/sec	1.097	km/hr
feet/sec	0.5921	knots
feet/sec	18.29	meters/min
feet/sec	0.6818	miles/hr
feet/sec	0.01136	miles/min
feet/sec/sec	30.48	cm/sec/sec
feet/sec/sec	1.097	km/hr/sec
feet/sec/sec	0.3048	meters/sec/sec
feet/sec/sec	0.6818	miles/hr/sec
feet/100 feet	1.0	percent grade
firkins (British)	9	gallons (British)
firkins (U.S.)	9	gallons (U.S.)
foot-candle (ft-c)	10.764	lumen/sq m
foot-poundals	3.9951×10^{-5}	BTU (mean)
foot-poundals	0.0421420	joules (abs)
foot-pounds	0.0012854	BTU (mean)
foot-pounds	0.32389	calories, gram (mean)
foot-pounds	1.13558×10^{7}	ergs
foot-pounds	32.174	foot-poundals
foot-pounds	5.050×10^{-7}	hp-hr
foot-pounds	1.35582	joules (abs)
foot-pounds	3.241×10^{-4}	kilogram-calories
foot-pounds	0.138255	kilogram-meters
foot-pounds	3.766×10^{-7}	kwh
foot-pounds	0.013381	liter-atmospheres
foot-pounds	3.7662×10^{-4}	watt-hours (abs)
foot-pounds/minute	1.286×10^{-3}	BTU/minute
foot-pounds/minute	0.01667	foot-pounds/sec
foot-pounds/minute	3.030×10^{-5}	hp
foot-pounds/minute	3.241×10^{-4}	kg-calories/min
foot-pounds/minute	2.260×10^{-5}	kw
foot-pounds/second	4.6275	BTU (mean)/hour
foot-pounds/second	0.07717	BTU/minute
foot-pounds/second	0.0018182	horsepower
foot-pounds/second	0.01945	kg-calories/min
foot-pounds/second	0.001356	kilowatts
foot-pounds/second	1.35582	watts (abs)
furlongs	660.0	feet
furlongs	201.17	meters
furlongs	0.125	miles (U.S.)
furlongs	40.0	rods
gallons (Br.)	3.8125×10^{-2}	barrels (U.S.)

Multiply	*by*	*to obtain*
gallons (Br.)	4516.086	cubic centimeters
gallons (Br.)	0.16053	cu ft
gallons (Br.)	277.4	cu inches
gallons (Br.)	1230	drams (U.S. fluid)
gallons (Br.)	4.54596	liters
gallons (Br.)	7.9620×10^4	minims (Br.)
gallons (Br.)	7.3783×10^4	minims (U.S.)
gallons (Br.)	4545.96	mL
gallons (Br.)	1.20094	gallons (U.S.)
gallons (Br.)	160	ounces (Br., fl.)
gallons (Br.)	153.72	ounces (U.S., fl.)
gallons (Br.)	10	pounds (avoirdupois) of water at 62°F
gallons (U.S.)	3.068×10^{-4}	acre-ft
gallons (U.S.)	0.031746	barrels (U.S.)
gallons (U.S.)	3785.434	cubic centimeters
gallons (U.S.)	0.13368	cubic feet (U.S.)
gallons (U.S.)	231	cubic inches
gallons (U.S.)	3.785×10^{-3}	cubic meters
gallons (U.S.)	4.951×10^{-3}	cubic yards
gallons (U.S.)	1024	drams (U.S., fluid)
gallons (U.S.)	0.83268	gallons (Br.)
gallons (U.S.)	0.83267	imperial gal
gallons (U.S.)	3.78533	liters
gallons (U.S.)	6.3950×10^4	minims (Br.)
gallons (U.S.)	6.1440×10^4	minims (U.S.)
gallons (U.S.)	3785	mL
gallons (U.S.)	133.23	ounces (Br., fluid)
gallons (U.S.)	128	ounces (U.S., fluid)
gallons	8	pints (liq.)
gallons	4	quarts (liq.)
gal water (U.S.)	8.345	lb of water
gallons/acre	0.00935	cu m/ha
gallons/day	4.381×10^{-5}	liters/sec
gpd/acre	0.00935	cu m/day/ha
gpd/acre	9.353	liter/day/ha
gallons/capita/day	3.785	liters/capita/day
gpd/cu yd	5.0	L/day/cu m
gpd/ft	0.01242	cu m/day/m
gpd/sq ft	0.0408	cu m/day/sq m
gpd/sq ft	1.698×10^{-5}	cubic meters/hour/sq meter
gpd/sq ft	0.283	cu meter/minute/ha
gpm (gal/min)	8.0208	cfh (cu ft/hr)
gpm	2.228×10^{-3}	cfs (cu ft/sec)

Multiply	*by*	*to obtain*
gpm	4.4021	cubic meters/hr
gpm	0.00144	MGD
gpm	0.0631	liters/sec
gpm/sq ft	2.445	cu meters/hour/sq meter
gpm/sq ft	40.7	L/min/sq meter
gpm/sq ft	0.679	liter/sec/sq meter
gallons/sq ft	40.743	liters/sq meter
gausses (abs)	3.3358×10^{-4}	electrostatic cgs units of magnetic flux density
gausses (abs)	0.99966	gausses (int)
gausses (abs)	1	lines/square centimeter
gausses (abs)	6.452	lines/sq in
gausses (abs)	1	maxwells (abs)/square centimeters
gausses (abs)	6.4516	maxwells (abs)/square inch
gausses (abs)	10^{-8}	webers/sq cm
gausses (abs)	6.452×10^{-8}	webers/sq in
gausses (abs)	10^{-4}	webers/sq meter
gilberts (abs)	0.07958	abampere turns
gilberts (abs)	0.7958	ampere turns
gilberts (abs)	2.998×10^{10}	electrostatic cgs units of magneto motive force
gilberts/cm	0.7958	amp-turns/cm
gilberts/cm	2.021	amp-turns/in
gilberts/cm	79.58	amp-turns/meter
gills (Br.)	142.07	cubic cm
gills (Br.)	5	ounces (British, fluid)
gills (U.S.)	32	drams (fluid)
gills	0.1183	liters
gills	0.25	pints (liq.)
grade	0.01571	radian
grains	0.036571	drams (avoirdupois)
grains	0.01667	drams (troy)
grains (troy)	1.216	grains (avdp)
grains (troy)	0.06480	grams
grains (troy)	6.480×10^{-5}	kilograms
grains (troy)	64.799	milligrams
grains (troy)	2.286×10^{-3}	ounces (avdp)
grains (troy)	2.0833×10^{-3}	ounces (troy)
grains (troy)	0.04167	pennyweights (troy)
grains	1/7000	pounds (avoirdupois)
grains	1.736×10^{-4}	pounds (troy)
grains	6.377×10^{-8}	tons (long)
grains	7.142×10^{-8}	tons (short)
grains/imp gal	14.254	mg/L

Multiply	*by*	*to obtain*
grains/imp. gal	14.254	parts/million (ppm)
grains/U.S. gal	17.118	mg/L
grains/U.S. gal	17.118	parts/million (ppm)
grains/U.S. gal	142.86	lb/mil gal
grams	0.5611	drams (avdp)
grams	0.25721	drams (troy)
grams	980.7	dynes
grams	15.43	grains
grams	9.807×10^{-5}	joules/cm
grams	9.807×10^{-3}	joules/meter (newtons)
grams	10^{-3}	kilograms
grams	10^{3}	milligrams
grams	0.0353	ounces (avdp)
grams	0.03215	ounces (troy)
grams	0.07093	poundals
grams	2.205×10^{-3}	pounds
grams	2.679×10^{-3}	pounds (troy)
grams	9.842×10^{-7}	tons (long)
grams	1.102×10^{-6}	tons (short)
grams-calories	4.1868×10^{7}	ergs
gram-calories	3.0880	foot-pounds
gram-calories	1.5597×10^{-6}	horsepower-hr
gram-calories	1.1630×10^{-6}	kilowatt-hr
gram-calories	1.1630×10^{-3}	watt-hr
gram-calories	3.968×10^{-3}	British Thermal Units (BTU)
gram-calories/sec	14.286	BTU/hr
gram-centimeters	9.2967×10^{-8}	BTU (mean)
gram-centimeters	2.3427×10^{-5}	calories, gram (mean)
gram-centimeters	980.7	ergs
gram-centimeters	7.2330×10^{-5}	foot-pounds
gram-centimeters	9.8067×10^{-5}	joules (abs)
gram-centimeters	2.344×10^{-8}	kilogram-calories
gram-centimeters	10^{-5}	kilogram-meters
gram-centimeters	2.7241×10^{-8}	watt-hours
grams-centimeters^2 (moment of inertia)	2.37305×10^{-6}	pounds-feet^2
grams-centimeters^2 (moment of inertia)	3.4172×10^{-4}	pounds-inch^2
gram-centimeters/second	1.3151×10^{-7}	hp
gram-centimeters/second	9.8067×10^{-8}	kilowatts
gram-centimeters/second	0.065552	lumens
gram-centimeters/second	9.80665×10^{-5}	watt (abs)
grams/cm	5.600×10^{-3}	pounds/inch
grams/cu cm	62.428	pounds/cubic foot
grams/cu cm	0.03613	pounds/cubic inch

Multiply	*by*	*to obtain*
grams/cu cm	8.3454	pounds/gallon (U.S.)
grams/cu cm	3.405×10^{-7}	pounds/mil-foot
grams/cu ft	35.314	grams/cu meter
grams/cu ft	10^6	micrograms/cu ft
grams/cu ft	35.314×10^6	micrograms/cu meter
grams/cu ft	35.3145×10^3	milligrams/cu meter
grams/cu ft	2.2046	pounds/1000 cu ft
grams/cu m	0.43700	grains/cubic foot
grams/cu m	0.02832	grams/cu ft
grams/cu m	28.317×10^3	micrograms/cu ft
grams/cu m	0.06243	pounds/cu ft
grams/liter	58.417	grains/gallon (U.S.)
grams/liter	9.99973×10^{-4}	grams/cubic centimeter
grams/liter	1000	mg/L
grams/liter	1000	parts per million (ppm)
grams/liter	0.06243	pounds/cubic foot
grams/liter	8.345	lb/1000 gal
grams/sq centimeter	2.0481	pounds/sq ft
grams/sq centimeter	0.0142234	pounds/square inch
grams/sq ft	10.764	grams/sq meter
grams/sq ft	10.764×10^3	kilograms/sq km
grams/sq ft	1.0764	milligrams/sq cm
grams/sq ft	10.764×10^3	milligrams/sq meter
grams/sq ft	96.154	pounds/acre
grams/sq ft	2.204	pounds/1000 sq ft
grams/sq ft	30.73	tons/sq mile
grams/sq meter	0.0929	grams/sq ft
grams/sq meter	1000	kilograms/sq km
grams/sq meter	0.1	milligrams/square cm
grams/sq meter	1000	milligrams/sq meter
grams/sq meter	8.921	pounds/acre
grams/sq meter	0.2048	pounds/1000 sq ft
grams/sq meter	2.855	tons/sq mile
g (gravity)	9.80665	meters/sec^2
g (gravity)	32.174	ft/sec^2
hand	10.16	cm
hands	4	inches
hectare (ha)	2.471	acre
hectares	1.076×10^5	sq feet
hectograms	100	grams
hectoliters	100	liters
hectometers	100	meters
hectowatts	100	watts
hemispheres	0.5	spheres

Multiply	*by*	*to obtain*
hemispheres	4	spherical right angles
hemispheres	6.2832	steradians
henries (abs)	10^9	abhenries
henries	1000.0	millihenries
henries (abs)	1.1126×10^{-12}	stathenries
hogsheads (British)	63	gallons (British)
hogsheads (British)	10.114	cubic feet
hogsheads (U.S.)	8.422	cubic feet
hogsheads (U.S.)	0.2385	cubic meters
hogsheads (U.S.)	63	gallons (U.S.)
horsepower	2545.08	BTU (mean)/hour
horsepower	42.44	BTU/min
horsepower	7.457×10^9	erg/sec
horsepower	33,000	ft lb/min
horsepower	550	foot-pounds/second
horsepower	7.6042×10^6	g cm/sec
horsepower, electrical	1.0004	horsepower
horsepower	10.70	kg.-calories/min
horsepower	0.74570	kilowatts (g = 980.665)
horsepower	498129	lumens
horsepower, continental	736	watts (abs)
horsepower, electrical	746	watts (abs)
horsepower (boiler)	9.803	kw
horsepower (boiler)	33.479	BTU/hr
horsepower-hours	2545	BTU (mean)
horsepower-hours	2.6845×10^{13}	ergs
horsepower-hours	6.3705×10^7	ft poundals
horsepower-hours	1.98×10^6	foot-pounds
horsepower-hours	641,190	gram-calories
horsepower-hours	2.684×10^6	joules
horsepower-hours	641.7	kilogram-calories
horsepower-hours	2.737×10^5	kilogram-meters
horsepower-hours	0.7457	kilowatt-hours (abs)
horsepower-hours	26,494	liter atmospheres (normal)
horsepower-hours	745.7	watt-hours
hours	4.167×10^{-2}	days
hours	60	minutes
hours	3600	seconds
hours	5.952×10^{-3}	weeks
hundredweights (long)	112	pounds
hundredweights (long)	0.05	tons (long)
hundredweights (short)	1600	ounces (avoirdupois)
hundredweights (short)	100	pounds
hundredweights (short)	0.0453592	tons (metric)

Multiply	*by*	*to obtain*
hundredweights (short)	0.0446429	tons (long)
inches (British)	2.540	centimeters
inches (U.S.)	2.54000508	centimeters
inches (British)	0.9999972	inches (U.S.)
inches	2.540×10^{-2}	meters
inches	1.578×10^{-5}	miles
inches	25.40	millimeters
inches	10^3	mils
inches	2.778×10^{-2}	yards
inches^2	6.4516×10^{-4}	meter^2
inches^3	1.6387×10^{-5}	meter^3
in. of mercury	0.0334	atm
in. of mercury	1.133	ft of water
in. of mercury (0°C)	13.609	inches of water (60°F)
in. of mercury	0.0345	kgs/square cm
in. of mercury at 32°F	345.31	kilograms/square meter
in. of mercury	33.35	millibars
in. of mercury	25.40	millimeters of mercury
in. of mercury (60°F)	3376.85	newton/meter^2
in. of mercury	70.73	pounds/square ft
in. of mercury at 32°F	0.4912	pounds/square inch
in. of water	0.002458	atmospheres
in. of water	0.0736	in. of mercury
in. of water (at 4°C)	2.540×10^{-3}	kgs/sq cm
in. of water	25.40	kgs/square meter
in. of water (60°F)	1.8663	millimeters of mercury (0°C)
in. of water (60°F)	248.84	newton/meter^2
in. of water	0.5781	ounces/square in
in. of water	5.204	pounds/square ft
in. of water	0.0361	psi
inches/hour	2.54	cm/hr
international ampere	.9998	ampere (absolute)
international volt	1.0003	volts (absolute)
international volt	1.593×10^{-19}	joules (absolute)
international volt	9.654×10^4	joules
joules	9.480×10^{-4}	BTU
joules (abs)	10^7	ergs
joules	23.730	foot poundals
joules (abs)	0.73756	foot-pounds
joules	3.7251×10^{-7}	horsepower hours
joules	2.389×10^{-4}	kg-calories
joules (abs)	0.101972	kilogram-meters
joules	9.8689×10^{-3}	liter atmospheres (normal)
joules	2.778×10^{-4}	watt-hrs

Multiply	*by*	*to obtain*
joules-sec	1.5258×10^{33}	quanta
joules/cm	1.020×10^{4}	grams
joules/cm	10^{7}	dynes
joules/cm	100.0	joules/meter (newtons)
joules/cm	723.3	poundals
joules/cm	22.48	pounds
joules/liter	0.02681	BTU/cu ft
joules/m^2-sec	0.3167	BTU/ft^2-hr
joules/sec	3.41304	BTU/hr
joules/sec	0.056884	BTU/min
joules/sec	1×10^{7}	erg/sec
joules/sec	44.254	ft lb/min
joules/sec	0.73756	ft lb/sec
joules/sec	1.0197×10^{4}	g cm/sec
joules/sec	1.341×10^{-3}	hp
joules/sec	0.01433	kg cal/min
joules/sec	0.001	kilowatts
joules/sec	668	lumens
joules/sec	1	watts
kilograms	564.38	drams (avdp)
kilograms	257.21	drams (troy)
kilograms	980,665	dynes
kilograms	15,432	grains
kilograms	1000	grams
kilograms	0.09807	joules/cm
kilograms	9.807	joules/meter (newtons)
kilograms	1×10^{6}	milligrams
kilograms	35.274	ounces (avdp)
kilograms	32.151	ounces (troy)
kilograms	70.93	poundals
kilograms	2.20462	pounds (avdp)
kilograms	2.6792	pounds (troy)
kilograms	9.84207×10^{-4}	tons (long)
kilograms	0.001	tons (metric)
kilograms	0.0011023	tons (short)
kilogram-calories	3.968	British Thermal Units (BTU)
kilogram-calories	3086	foot-pounds
kilogram-calories	1.558×10^{-3}	horsepower-hours
kilogram-calories	4186	joules
kilogram-calories	426.6	kilogram-meters
kilogram-calories	4.186	kilojoules
kilogram-calories	1.162×10^{-3}	kilowatt-hours
kg-cal/min	238.11	BTU/hr
kg-cal/min	3.9685	BTU/min

Multiply	*by*	*to obtain*
kg-cal/min	6.9770×10^{8}	erg/sec
kg-cal/min	3087.4	ft-lb/min
kg-cal/min	51.457	ft-lb/sec
kg-cal/min	7.1146×10^{5}	g cm/sec
kg-cal/min	0.0936	hp
kg-cal/min	69.769	joules/sec
kg-cal/min	0.0698	kw
kg-cal/min	46636	lumens
kg-cal/min	69.767	watts
kgs-cms. squared	2.373×10^{-3}	pounds-feet squared
kgs-cms. squared	0.3417	pounds-inches squared
kilogram-force (kgf)	9.80665	newton
kilogram-meters	0.0092967	BTU (mean)
kilogram-meters	2.3427	calories, gram (mean)
kilogram-meters	9.80665×10^{7}	ergs
kilogram-meters	232.71	ft poundals
kilogram-meters	7.2330	foot-pounds
kilogram-meters	3.6529×10^{-6}	horsepower-hours
kilogram-meters	9.80665	joules (abs)
kilogram-meters	2.344×10^{-3}	kilogram-calories
kilogram-meters	2.52407×10^{-6}	kilowatt-hours (abs)
kilogram-meters	2.7241×10^{-6}	kilowatt-hours
kilogram-meters	0.096781	liter atmospheres (normal)
kilogram-meters	6.392×10^{-7}	pounds carbon to CO_2
kilogram-meters	9.579×10^{-6}	pounds water evap. at 212°F
kilograms/cubic meter	10^{-3}	grams/cubic cm
kilograms/cubic meter	0.06243	pounds/cubic foot
kilograms/cubic meter	3.613×10^{-5}	pounds/cubic inch
kilograms/cubic meter	3.405×10^{-10}	pounds/mil. foot
kilograms/m^3-day	0.0624	lb/cu ft-day
kilograms/cu meter-day	62.43	pounds/1000 cu ft-day
kilograms/ha	0.8921	pounds/acre
kilograms/meter	0.6720	pounds/foot
kilograms/sq cm	980,665	dynes
kilograms/sq cm	0.96784	atmosphere
kilograms/sq cm	32.81	feet of water
kilograms/sq cm	28.96	inches of mercury
kilograms/sq cm	735.56	mm of mercury
kilograms/sq cm	2048	pounds/sq ft
kilograms/sq cm	14.22	pounds/square inch
kilograms/sq km	92.9×10^{-6}	grams/sq ft
kilograms/sq km	0.001	grams/sq meter
kilograms/sq km	0.0001	milligrams/sq cm
kilograms/sq km	1.0	milligrams/sq meter

Multiply	*by*	*to obtain*
kilograms/sq km	8.921×10^{-3}	pounds/acre
kilograms/sq km	204.8×10^{-6}	pounds/1000 sq ft
kilograms/sq km	2.855×10^{-3}	tons/sq mile
kilograms/sq meter	9.6784×10^{-5}	atmospheres
kilograms/sq meter	98.07×10^{-6}	bars
kilograms/sq meter	98.0665	dynes/sq centimeters
kilograms/sq meter	3.281×10^{-3}	feet of water at 39.2°F
kilograms/sq meter	0.1	grams/sq centimeters
kilograms/sq meter	2.896×10^{-3}	inches of mercury at 32°F
kilograms/sq meter	0.07356	mm of mercury at 0°C
kilograms/sq meter	0.2048	pounds/square foot
kilograms/sq meter	0.00142234	pounds/square inch
kilograms/sq mm.	10^6	kg/square meter
kilojoule	0.947	BTU
kilojoules/kilogram	0.4295	BTU/pound
kilolines	1000.0	maxwells
kiloliters	10^3	liters
kilometers	10^5	centimeters
kilometers	3281	feet
kilometers	3.937×10^4	inches
kilometers	10^3	meters
kilometers	0.53961	miles (nautical)
kilometers	0.6214	miles (statute)
kilometers	10^6	millimeters
kilometers	1093.6	yards
kilometers/hr	27.78	cm/sec
kilometers/hr	54.68	feet/minute
kilometers/hr	0.9113	ft/sec
kilometers/hr	0.5396	knot
kilometers/hr	16.67	meters/minute
kilometers/hr	0.2778	meters/sec
kilometers/hr	0.6214	miles/hour
kilometers/hour/sec	27.78	cms/sec/sec
kilometers/hour/sec	0.9113	ft/sec/sec
kilometers/hour/sec	0.2778	meters/sec/sec
kilometers/hour/sec	0.6214	miles/hr/sec
kilometers/min	60	kilometers/hour
kilonewtons/sq m	0.145	psi
kilowatts	56.88	BTU/min
kilowatts	4.425×10^4	foot-pounds/min
kilowatts	737.6	ft-lb/sec
kilowatts	1.341	horsepower
kilowatts	14.34	kg-cal/min
kilowatts	10^3	watts

Multiply	*by*	*to obtain*
kilowatt-hrs	3413	BTU (mean)
kilowatt-hrs	3.600×10^{13}	ergs
kilowatt-hrs	2.6552×10^{6}	foot-pounds
kilowatt-hrs	859,850	gram-calories
kilowatt-hrs	1.341	horsepower hours
kilowatt-hrs	3.6×10^{6}	joules
kilowatt-hrs	860.5	kg-calories
kilowatt-hrs	3.6709×10^{5}	kilogram-meters
kilowatt-hrs	3.53	pounds of water evaporated from and at 212°F
kilowatt-hrs	22.75	pounds of water raised from 62° to 212°F
knots	6080	feet/hr
knots	1.689	feet/sec
knots	1.8532	kilometers/hr
knots	0.5144	meters/sec
knots	1.0	miles (nautical)/hour
knots	1.151	miles (statute)/hour
knots	2,027	yards/hr
lambert	2.054	candle/in^2
lambert	929	footlambert
lambert	0.3183	stilb
langley	1	15° gram-calorie/cm^2
langley	3.6855	BTU/ft^2
langley	0.011624	Int. kw-hr/m^2
langley	4.1855	joules (abs)/cm^2
leagues (nautical)	3	miles (nautical)
leagues (statute)	3	miles (statute)
light years	63,274	astronomical units
light years	9.4599×10^{12}	kilometers
light years	5.8781×10^{12}	miles
lignes (Paris lines)	1/12	ponces (Paris inches)
lines/sq cm	1.0	gausses
lines/sq in	0.1550	gausses
lines/sq in	1.550×10^{-9}	webers/sq cm
lines/sq in	10^{-8}	webers/sq in
lines/sq in	1.550×10^{-5}	webers/sq meter
links (engineer's)	12.0	inches
links (Gunter's)	0.01	chains (Gunter's)
links (Gunter's)	0.66	feet
links (Ramden's)	0.01	chains (Ramden's)
links (Ramden's)	1	feet
links (surveyor's)	7.92	inches
liters	8.387×10^{-3}	barrels (U.S.)

Multiply	*by*	*to obtain*
liters	0.02838	bushels (U.S. dry)
liters	1000.028	cubic centimeters
liters	0.035316	cubic feet
liters	61.025	cu inches
liters	10^{-3}	cubic meters
liters	1.308×10^{-3}	cubic yards
liters	270.5179	drams (U.S. fl)
liters	0.21998	gallons (Br.)
liters	0.26417762	gallons (U.S.)
liters	16,894	minims (Br.)
liters	16,231	minims (U.S.)
liters	35.196	ounces (Br. fl)
liters	33.8147	ounces (U.S. fl)
liters	2.113	pints (liq.)
liters	1.0566828	quarts (U.S. liq.)
liter-atmospheres (normal)	0.096064	BTU (mean)
liter-atmospheres (normal)	24.206	calories, gram (mean)
liter-atmospheres (normal)	1.0133×10^{9}	ergs
liter-atmospheres (normal)	74.735	foot-pounds
liter-atmospheres (normal)	3.7745×10^{-5}	horsepower hours
liter-atmospheres (normal)	101.33	joules (abs)
liter-atmospheres (normal)	10.33	kilogram-meters
liter-atmospheres (normal)	2.4206×10^{-2}	kilogram calories
liter-atmospheres (normal)	2.815×10^{-5}	kilowatt-hours
liter/cu m-sec	60.0	cfm/1000 cu ft
liters/minute	5.885×10^{-4}	cubic feet/sec
liters/minute	4.403×10^{-3}	gallons/sec
liter/person-day	0.264	gpcd
liters/sec	2.119	cu ft /min
liters/sec	3.5316×10^{-2}	cu ft /sec
liters/sec	15.85	gallons/minute
liters/sec	0.02282	MGD
$\log_{10} N$	2.303	$\log_e N$ or ln N
$\log_e N$ or ln N	0.4343	$\log_{10} N$
lumens	0.07958	candle-power (spherical)
lumens	0.00147	watts of maximum visibility radiation
lumens/sq. centimeters	1	lamberts
lumens/sq cm/steradian	3.1416	lamberts
lumens/sq ft	1	foot-candles
lumens/sq ft	10.764	lumens/sq meter
lumens/sq ft/steradian	3.3816	millilamberts
lumens/sq meter	0.09290	foot-candles or lumens/sq
lumens/sq meter	10^{-4}	phots
lux	0.09290	foot-candles

Multiply	*by*	*to obtain*
lux	1	lumens/sq meter
lux	10^{-4}	phots
maxwells	0.001	kilolines
maxwells	10^{-8}	webers
megajoule	0.3725	horsepower-hour
megalines	10^{6}	maxwells
megohms	10^{12}	microhms
megohms	10^{6}	ohms
meters	10^{10}	angstrom units
meters	100	centimeters
meters	0.5467	fathoms
meters	3.280833	feet (U.S.)
meters	39.37	inches
meters	10^{-3}	kilometers
meters	5.396×10^{-4}	miles (naut.)
meters	6.2137×10^{-4}	miles (statute)
meters	10^{3}	millimeters
meters	10^{9}	millimicrons
meters	1.09361	yards (U.S.)
meters	1.179	varas
meter-candles	1	lumens/sq meter
meter-kilograms	9.807×10^{7}	centimeter-dynes
meter-kilograms	10^{5}	centimeter-grams
meter-kilograms	7.233	pound-feet
meters/minute	1.667	centimeters/sec
meters/minute	3.281	feet/minute
meters/minute	0.05468	feet/second
meters/minute	0.06	kilograms/hour
meters/minute	0.03238	knots
meters/minute	0.03728	miles/hour
meters/second	196.8	feet/minute
meters/second	3.281	feet/second
meters/second	3.6	kilometers/hour
meters/second	0.06	kilometers/min
meters/second	1.944	knots
meters/second	2.23693	miles/hour
meters/second	0.03728	miles/minute
meters/sec/sec	100.0	cm/sec/sec
meters/sec/sec	3.281	feet/sec/sec
meters/sec/sec	3.6	km/hour/sec
meters/sec/sec	2.237	miles/hour/sec
microfarad	10^{-6}	farads
micrograms	10^{-6}	grams
micrograms/cu ft	10^{-6}	grams/cu ft

Multiply	*by*	*to obtain*
micrograms/cu ft	35.314×10^{-6}	grams/cu m
micrograms/cu ft	35.314	microgram/cu m
micrograms/cu ft	35.314×10^{-3}	milligrams/cu m
micrograms/cu ft	2.2046×10^{-6}	pounds/1000 cu ft
micrograms/cu m	28.317×10^{-9}	grams/cu ft
micrograms/cu m	10^{-6}	grams/ cu m
micrograms/cu m	0.02832	micrograms/cu ft
micrograms/cu m	0.001	milligrams/cu m
micrograms/cu m	62.43×10^{-9}	pounds/1000 cu ft
micrograms/cu m	$\frac{0.02404}{\text{molecular weight of gas}}$	ppm by volume (20°C)
micrograms/cu m	834.7×10^{-6}	ppm by weight
micrograms/liter	1000.0	micrograms/cu m
micrograms/liter	1.0	milligrams/cu m
micrograms/liter	62.43×10^{-9}	pounds/cu ft
micrograms/liter	$\frac{24.04}{\text{molecular weight of gas}}$	ppm by volume (20°C)
micrograms/liter	0.834.7	ppm by weight
microhms	10^{-12}	megohms
microhms	10^{-6}	ohms
microliters	10^{-6}	liters
microns	10^{4}	angstrom units
microns	1×10^{-4}	centimeters
microns	3.9370×10^{-5}	inches
microns	10^{-6}	meters
miles (naut.)	6080.27	feet
miles (naut.)	1.853	kilometers
miles (naut.)	1.853	meters
miles (naut.)	1.1516	miles (statute)
miles (naut.)	2027	yards
miles (statute)	1.609×10^{5}	centimeters
miles (statute)	5280	feet
miles (statute)	6.336×10^{4}	inches
miles (statute)	1.609	kilometers
miles (statute)	1609	meters
miles (statute)	0.8684	miles (naut.)
miles (statute)	320	rods
miles (statute)	1760	yards
miles/hour	44.7041	centimeter/second
miles/hour	88	feet/min
miles/hour	1.4667	feet/sec
miles/hour	1.6093	kilometers/hour
miles/hour	0.02682	km/min

Multiply	*by*	*to obtain*
miles/hour	0.86839	knots
miles/hour	26.82	meters/min
miles/hour	0.447	meters/sec
miles/hour	0.1667	miles/min
miles/hour/sec	44.70	cms/sec/sec
miles/hour/sec	1.4667	ft/sec/sec
miles/hour/sec	1.6093	km/hour/sec
miles/hour/sec	0.4470	m/sec/sec
miles/min	2682	centimeters/sec
miles/min	88	ft/sec
miles/min	1.609	km/min
miles/min	0.8684	knots/min
miles/min	60	miles/hour
miles-feet	9.425×10^{-6}	cu inches
millibars	0.00987	atmospheres
millibars	0.30	inches of mercury
millibars	0.75	millimeters of mercury
milliers	10^3	kilograms
millimicrons	1×10^{-9}	meters
milligrams	0.01543236	grains
milligrams	10^{-3}	grams
milligrams	10^{-6}	kilograms
milligrams	3.5274×10^{-5}	ounces (avdp)
milligrams	2.2046×10^{-6}	pounds (avdp)
milligrams/assay ton	1	ounces (troy)/ton (short)
milligrams/cu m	283.2×10^{-6}	grams/cu ft
milligrams/cu m	0.001	grams/cu m
milligrams/cu m	1000.0	micrograms/cu m
milligrams/cu m	28.32	micrograms/cu ft
milligrams/cu m	1.0	micrograms/liter
milligrams/cu m	62.43×10^{-6}	pounds/1000 cu ft
milligrams/cu m	$\frac{24.04}{\text{molecular weight of gas}}$	ppm by volume (20°C)
milligrams/cu m	0.8347	ppm by weight
milligrams/joule	5.918	pounds/horsepower-hour
milligrams/liter	0.05841	grains/gallon
milligrams/liter	0.07016	grains/imp. gal
milligrams/liter	0.0584	grains/U.S. gal
milligrams/liter	1.0	parts/million
milligrams/liter	8.345	lb/mil gal
milligrams/sq cm	0.929	grams/sq ft
milligrams/sq cm	10.0	grams/sq meter
milligrams/sq cm	10^4	kilograms/sq km

Multiply	*by*	*to obtain*
milligrams/sq cm	10^4	milligrams/sq meter
milligrams/sq cm	2.048	pounds/1000 sq ft
milligrams/sq cm	89.21	pounds/acre
milligrams/sq cm	28.55	tons/sq mile
milligrams/sq meter	92.9×10^{-6}	grams/sq ft
milligrams/sq meter	0.001	grams/sq meter
milligrams/sq meter	1.0	kilograms/sq km
milligrams/sq meter	0.0001	milligrams/sq cm
milligrams/sq meter	8.921×10^{-3}	pounds/acre
milligrams/sq meter	204.8×10^{-6}	pounds/1000 sq ft
milligrams/sq meter	2.855×10^{-3}	tons/sq mile
millihenries	0.001	henries
milliliters	1	cubic centimeters
milliliters	3.531×10^{-5}	cu ft
milliliters	6.102×10^{-2}	cu in
milliliters	10^{-6}	cu m
milliliters	2.642×10^{-4}	gal (U.S.)
milliliters	10^{-3}	liters
milliliters	0.03381	ounces (U.S. fl)
millimeters	0.1	centimeters
millimeters	3.281×10^{-3}	feet
millimeters	0.03937	inches
millimeters	10^{-6}	kilometers
millimeters	0.001	meters
millimeters	6.214×10^{-7}	miles
millimeters	39.37	mils
millimeters	1.094×10^{-3}	yards
millimeters of mercury	1.316×10^{-3}	atmospheres
millimeters of mercury	0.0394	inches of mercury
millimeters of mercury (0°C)	0.5358	inches of water (60°F)
millimeters of mercury	1.3595×10^{-3}	kg/sq cm
millimeter of mercury (0°C)	133.3224	newton/meter^2
millimeters of mercury	0.01934	pounds/sq in
millimeters/sec	11.81	feet/hour
million gallons	306.89	acre-ft
million gallons	3785.0	cubic meters
million gallons	3.785	mega liters (1×10^6)
million gallons/day (MGD)	1.547	cu ft/sec
MGD	3785	cu m/day
MGD	0.0438	cubic meters/sec
MGD	43.808	liters/sec
MGD/acre	9360	cu m/day/ha
MGD/acre	0.039	cu meters/hour/sq meter

Multiply	*by*	*to obtain*
mils	0.002540	centimeters
mils	8.333×10^{-5}	feet
mils	0.001	inches
mils	2.540×10^{-8}	kilometers
mils	25.40	microns
mils	2.778×10^{-5}	yards
miner's in.	1.5	cu ft/min
miner's inches (Ariz., Calif. Mont., and Ore.)	0.025	cubic feet/second
miner's in. (Colorado)	0.02604	cubic feet/second
miner's inches (Idaho, Kan., Neb., Nev., N. Mex., N. Dak., S. Dak. and Utah)	0.020	cubic feet/second
minims (British)	0.05919	cubic centimeter
minims (U.S.)	0.06161	cubic centimeters
minutes (angles)	0.01667	degrees
minutes (angles)	1.852×10^{-4}	quadrants
minutes (angles)	2.909×10^{-4}	radians
minutes (angle)	60	seconds (angle)
months (mean calendar)	30.4202	days
months (mean calendar)	730.1	hours
months (mean calendar)	43805	minutes
months (mean calendar)	2.6283×10^{6}	seconds
myriagrams	10	kilograms
myriameters	10	kilometers
myriawatts	10	kilowatts
nepers	8.686	decibels
newtons	10^{5}	dynes
newtons	0.10197	kilograms
newtons	0.22481	pounds
newtons/sq meter	1.00	pascals (Pa)
noggins (British)	1/32	gallons (British)
No./cu.cm.	28.316×10^{3}	No./cu ft
No./cu.cm.	10^{6}	No./cu meter
No./cu.cm.	1000.0	No./liter
No./cu.ft.	35.314×10^{-6}	No./cu cm
No./cu.ft.	35.314	No./cu meter
No./cu.ft.	35.314×10^{-3}	No./liter
No./cu. meter	10^{-6}	No./cu cm
No./cu. meter	28.317×10^{-3}	No./cu ft
No./cu. meter	0.001	No./liter
No./liter	0.001	No./cu cm
No./liter	28.316	No./cu ft

Multiply	*by*	*to obtain*
No./liter	1000.0	No./cu meter
oersteds (abs)	1	electromagnetic cgs units of magnetizing force
oersteds (abs)	2.9978×10^{10}	electrostatic cgs units of magnetizing force
ohms	10^{9}	abohms
ohms	1.1126×10^{-12}	statohms
ohms	10^{-6}	megohms
ohms	10^{6}	microhms
ohms (International)	1.0005	ohms (absolute)
ounces (avdp)	16	drams (avoirdupois)
ounces (avdp)	7.2917	drams (troy)
ounces (avdp)	437.5	grains
ounces (avdp)	28.349527	grams
ounces (avdp)	0.028350	kilograms
ounces (avdp)	2.8350×10^{4}	milligrams
ounces (avdp)	0.9114583	ounces (troy)
ounces (avdp)	0.0625	pounds (avoirdupois)
ounces (avdp)	0.075955	pounds (troy)
ounces (avdp)	2.790×10^{-5}	tons (long)
ounces (avdp)	2.835×10^{-5}	tons (metric)
ounces (avdp)	3.125×10^{-5}	tons (short)
ounces (Br. fl)	2.3828×10^{-4}	barrels (U.S.)
ounces (Br. fl)	1.0033×10^{-3}	cubic feet
ounces (Br. fl)	1.73457	cubic inches
ounces (Br. fl)	7.6860	drams (U.S. fl)
ounces (Br. fl)	6.250×10^{-3}	gallons (Br.)
ounces (Br. fl)	0.07506	gallons (U.S.)
ounces (Br. fl)	2.84121×10^{-2}	liters
ounces (Br. fl)	480	minims (Br.)
ounces (Br. fl)	461.160	minims (U.S.)
ounces (Br. fl)	28.4121	mL
ounces (Br. fl)	0.9607	ounces (U.S. fl)
ounces (troy)	17.554	drams (avdp)
ounces (troy)	8	drams (troy)
ounces (troy)	480	grains (troy)
ounces (troy)	31.103481	grams
ounces (troy)	0.03110	kilograms
ounces (troy)	1.09714	ounces (avoirdupois)
ounces (troy)	20	pennyweights (troy)
ounces (troy)	0.068571	pounds (avdp)
ounces (troy)	0.08333	pounds (troy)
ounces (troy)	3.061×10^{-5}	tons (long)
ounces (troy)	3.429×10^{-5}	tons (short)

Multiply	*by*	*to obtain*
ounces (U.S. fl)	2.48×10^{-4}	barrels (U.S.)
ounces (U.S. fl)	29.5737	cubic centimeters
ounces (U.S. fl)	1.0443×10^{-3}	cubic feet
ounces (U.S. fl)	1.80469	cubic inches
ounces (U.S. fl)	8	drams (fluid)
ounces (U.S. fl)	6.5053×10^{-3}	gallons (Br.)
ounces (U.S. fl)	7.8125×10^{-3}	gallons (U.S.)
ounces (U.S. fl)	29.5729	milliliters
ounces (U.S. fl)	499.61	minims (Br.)
ounces (U.S. fl)	480	minims (U.S.)
ounces (U.S. fl)	1.0409	ounces (Br. fl)
ounces/sq inch	4309	dynes/sq cm
ounces/sq. inch	0.0625	pounds/sq inch
paces	30	inches
palms (British)	3	inches
parsecs	3.260	light years
parsecs	3.084×10^{13}	kilometers
parsecs	3.084×10^{16}	meters
parsec	19×10^{12}	miles
parts/billion (ppb)	10^{-3}	mg/L
parts/million (ppm)	0.07016	grains/imp. gal.
parts/million	0.058417	grains/gallon (U.S.)
parts/million	1.0	mg/liter
parts/million	8.345	lbs/million gallons
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{24.04}$	micrograms/liter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{0.02404}$	micrograms/cu meter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{24.04}$	milligrams/cu meter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{28.8}$	ppm by weight
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{385.1 \times 10^{6}}$	pounds/cu ft
ppm by weight	1.198×10^{-3}	micrograms/cu meter
ppm by weight	1.198	micrograms/liter
ppm by weight	1.198	milligrams/cu meter
ppm by weight	$\frac{28.8}{\text{molecular weight of gas}}$	ppm by volume (20°C)
ppm by weight	7.48×10^{-6}	pounds/cu ft
pecks (British)	0.25	bushels (British)
pecks (British)	554.6	cubic inches

Multiply	*by*	*to obtain*
pecks (British)	9.091901	liters
pecks (U.S.)	0.25	bushels (U.S.)
pecks (U.S.)	537.605	cubic inches
pecks (U.S.)	8.809582	liters
pecks (U.S.)	8	quarts (dry)
pennyweights	24	grains
pennyweights	1.555174	grams
pennyweights	0.05	ounces (troy)
pennyweights (troy)	4.1667×10^{-3}	pounds (troy)
perches (masonry)	24.75	cubic feet
phots	929.0	foot-candles
phots	1	lumen incident/sq cm
phots	10^4	lux
picas (printers')	1/6	inches
pieds (French feet)	0.3249	meters
pints (dry)	33.6003	cubic inches
pints (liq.)	473.179	cubic centimeters
pints (liq.)	0.01671	cubic feet
pints (liq.)	4.732×10^{-4}	cubic meters
pints (liq.)	6.189×10^{-4}	cubic yards
pints (liq.)	0.125	gallons
pints (liq.)	0.4732	liters
pints (liq.)	16	ounces (U.S. fluid)
pints (liq.)	0.5	quarts (liq.)
planck's constant	6.6256×10^{-27}	erg-seconds
poise	1.00	gram/cm sec
poise	0.1	newton-second/meter^2
population equivalent (PE)	0.17	pounds BOD
pottles (British)	0.5	gallons (British)
pouces (Paris inches)	0.02707	meters
pouces (Paris inches)	0.08333	pieds (Paris feet)
poundals	13,826	dynes
poundals	14.0981	grams
poundals	1.383×10^{-3}	joules/cm
poundals	0.1383	joules/meter (newton)
poundals	0.01410	kilograms
poundals	0.031081	pounds
pounds (avdp)	256	drams (avdp)
pounds (avdp)	116.67	drams (troy)
pounds (avdp)	444,823	dynes
pounds (avdp)	7000	grains
pounds (avdp)	453.5924	grams
pounds (avdp)	0.04448	joules/cm
pounds (avdp)	4.448	joules/meter (newtons)

Multiply	*by*	*to obtain*
pounds (avdp)	0.454	kilograms
pounds (avdp)	4.5359×10^5	milligrams
pounds (avdp)	16	ounces (avdp)
pounds (avdp)	14.5833	ounces (troy)
pounds (avdp)	32.17	poundals
pounds (avdp)	1.2152778	pounds (troy)
pounds (avdp)	4.464×10^{-4}	tons (long)
pounds (avdp)	0.0005	tons (short)
pounds (troy)	210.65	drams (avdp)
pounds (troy)	96	drams (troy)
pounds (troy)	5760	grains
pounds (troy)	373.2418	grams
pounds (troy)	0.37324	kilograms
pounds (troy)	3.7324×10^5	milligrams
pounds (troy)	13.1657	ounces (avdp)
pounds (troy)	12.0	ounces (troy)
pounds (troy)	240.0	pennyweights (troy)
pounds (troy)	0.8229	pounds (avdp)
pounds (troy)	3.6735×10^{-4}	tons (long)
pounds (troy)	3.7324×10^{-4}	tons (metric)
pounds (troy)	4.1143×10^{-4}	tons (short)
pounds (avdp)-force	4.448	newtons
pounds-force-sec/ft^2	47.88026	newton-sec/$meter^2$
pounds (avdp)-mass	0.4536	kilograms
pounds-mass/ft^3	16.0185	kilogram/$meter^3$
pounds-mass/ft-sec	1.4882	mewton-sec/$meter^2$
pounds of BOD	5.882	population equivalent (PE)
pounds of carbon to CO_2	14,544	BTU (mean)
pounds of water	0.0160	cu ft
pounds of water	27.68	cu in
pounds of water	0.1198	gallons
pounds of water evaporated at 212°F	970.3	BTU
pounds of water per min	2.699×10^{-4}	cubic feet/sec
pound-feet	13,825	centimeter-grams
pound-feet (torque)	1.3558×10^7	dyne-centimeters
pound-feet	0.1383	meter-kilograms
pounds-feet squared	421.3	kg-cm squared
pounds-feet squared	144	pounds-inches squared
pounds-inches squared	2926	kg-cm squared
pounds-inches squared	6.945×10^{-3}	pounds-feet squared
pounds/acre	0.0104	grams/sq ft
pounds/acre	0.1121	grams/sq meter
pounds/acre	1.121	kg/ha
pounds/acre	112.1	kilograms/sq km

Multiply	*by*	*to obtain*
pounds/acre	0.01121	milligrams/sq cm
pounds/acre	112.1	milligrams/sq meter
pounds/acre	0.023	pounds/1000 sq ft
pounds/acre	0.32	tons/sq mile
pounds/acre/day	0.112	g/day/sq m
pounds/cu ft	0.0160	g/mL
pounds/cu ft	16.02	kg/cu m
pounds/cu ft	16.018×10^9	micrograms/cu meter
pounds/cu ft	16.018×10^6	micrograms/liter
pounds/cu ft	16.018×10^6	milligrams/cu meter
pounds/cu ft	$\frac{385.1 \times 10^6}{\text{molecular weight of gas}}$	ppm by volume (20°C)
pounds/cu ft	133.7×10^3	ppm by weight
pounds/cu ft	5.787×10^{-4}	lb/cu in
pounds/cu ft	5.456×10^{-9}	pounds/mil-foot
pounds/1000 cu ft	0.35314	grams/cu ft
pounds/1000 cu ft	16.018	grams/cu m
pounds/1000 cu ft	353.14×10^3	micrograms/cu ft
pounds/1000 cu ft	16.018×10^6	microgram/cu m
pounds/1000 cu ft	16.018×10^3	milligrams/cu m
pounds/cubic inch	27.68	grams/cubic cm
pounds/cubic inch	2.768×10^4	kgs/cubic meter
pounds/cubic inch	1728	pounds/cubic foot
pounds/cubic inch	9.425×10^{-6}	pounds/mil foot
pounds/day/acre-ft	3.68	g/day/cu m
pounds/day/cu ft	16	kg/day/cu m
pounds/day/cu yd	0.6	kg/day/cu m
pounds/day/sq ft	4,880	g/day/sq m
pounds/ft	1.488	kg/m
pounds/gal	$454\,\text{g}/3.7851\text{L} = 119.947$	g/liter
pounds/1000-gal	120	g/1000-liters
pounds/horsepower-hour	0.169	mg/joule
pounds/in	178.6	g/cm
pounds/mil-foot	2.306×10^6	gms/cu cm
pounds/mil gal	0.12	g/cu m
pounds/sq ft	4.725×10^{-4}	atmospheres
pounds/sq ft	0.01602	ft of water
pounds/sq ft	0.01414	inches of mercury
pounds/sq ft	4.8824×10^{-4}	kgs/sq cm
pounds/sq ft	4.88241	kilograms/square meter
pounds/sq ft	47.9	newtons/sq m
pounds/sq ft	6.944×10^{-3}	pounds/sq inch
pounds/1000 sq ft	0.4536	grams/sq ft

Multiply	*by*	*to obtain*
pounds/1000 sq ft	4.882	grams/sq meter
pounds/1000 sq ft	4882.4	kilograms/sq km
pounds/1000 sq ft	0.4882	milligrams/sq cm
pounds/1000 sq ft	4882.4	milligrams/sq meter
pounds/1000 sq ft	43.56	pounds/acre
pounds/1000 sq ft	13.94	tons/sq mile
pounds/sq in	0.068046	atmospheres
pounds/sq in	2.307	ft of water
pounds/sq in	70.307	grams/square centimeter
pounds/sq in	2.036	in of mercury
pounds/sq in	0.0703	kgs/square cm
pounds/sq in	703.07	kilograms/square meter
pounds/sq in	51.715	millimeters of mercury
pounds/sq in	6894.76	newton/meter^2
pounds/sq in	51.715	millimeters of mercury at 0°C
pounds/sq in	144	pounds/sq foot
pounds/sq in (abs)	1	pound/sq in (gage) + 14.696
proof (U.S.)	0.5	percent alcohol by volume
puncheons (British)	70	gallons (British)
quadrants (angle)	90	degrees
quadrants (angle)	5400	minutes
quadrants (angle)	3.24×10^5	seconds
quadrants (angle)	1.571	radians
quarts (dry)	67.20	cubic inches
quarts (liq.)	946.4	cubic centimeters
quarts (liq.)	0.033420	cubic feet
quarts (liq.)	57.75	cubic inches
quarts (liq.)	9.464×10^{-4}	cubic meters
quarts (liq.)	1.238×10^{-3}	cubic yards
quarts (liq.)	0.25	gallons
quarts (liq.)	0.9463	liters
quarts (liq.)	32	ounces (U.S., fl)
quarts (liq.)	0.832674	quarts (British)
quintals (long)	112	pounds
quintals (metric)	100	kilograms
quintals (short)	100	pounds
quires	24	sheets
radians	57.29578	degrees
radians	3438	minutes
radians	0.637	quadrants
radians	2.063×10^5	seconds
radians/second	57.30	degrees/second
radians/second	9.549	revolutions/min
radians/second	0.1592	revolutions/sec

Multiply	*by*	*to obtain*
radians/sec/sec	573.0	revs/min/min
radians/sec/sec	9.549	revs/min/sec
radians/sec/sec	0.1592	revs/sec/sec
reams	500	sheets
register tons (British)	100	cubic feet
revolutions	360	degrees
revolutions	4	quadrants
revolutions	6.283	radians
revolutions/minute	6	degrees/second
revolutions/minute	0.10472	radians/second
revolutions/minute	0.01667	revolutions/sec
revolutions/minute^2	0.0017453	radians/sec/sec
revs/min/min	0.01667	revs/min/sec
revs/min/min	2.778×10^{-4}	revs/sec/sec
revolutions/second	360	degrees/second
revolutions/second	6.283	radians/second
revolutions/second	60	revs/minute
revs/sec/sec	6.283	rads/sec/sec
revs/sec/sec	3600	revs/min/min
revs/sec/sec	60	revs/min/sec
reyns	6.8948×10^{6}	centipoises
rod	.25	chain (gunters)
rods	16.5	feet
rods	5.0292	meters
rods	3.125×10^{-3}	miles
rods (surveyors' means)	5.5	yards
roods (British)	0.25	acres
scruples	1/3	drams (troy)
scruples	20	grains
sections	1	square miles
seconds (mean solar)	1.1574×10^{-5}	days
seconds (angle)	2.778×10^{-4}	degrees
seconds (mean solar)	2.7778×10^{-4}	hours
seconds (angle)	0.01667	minutes
seconds (angle)	3.087×10^{-6}	quadrants
seconds (angle)	4.848×10^{-6}	radians
slugs	14.59	kilogram
slugs	32.174	pounds
space, entire (solid angle)	12.566	steradians
spans	9	inches
spheres (solid angle)	12.57	steradians
spherical right angles	0.25	hemispheres
spherical right angles	0.125	spheres
spherical right angles	1.571	steradians

Multiply	*by*	*to obtain*
square centimeters	1.973×10^{5}	circular mils
square centimeters	1.07639×10^{-3}	square feet (U.S.)
square centimeters	0.15499969	square inches (U.S.)
square centimeters	10^{-4}	square meters
square centimeters	3.861×10^{-11}	square miles
square centimeters	100	square millimeters
square centimeters	1.196×10^{-4}	square yards
square centimeters-square centimeter (moment of area)	0.024025	square inch-square inch
square chains (gunter's)	0.1	acres
square chains (gunter's)	404.7	square meters
square chains (Ramden's)	0.22956	acres
square chains (Ramden's)	10000	square feet
square feet	2.29×10^{-5}	acres
square feet	1.833×10^{8}	circular mils
square feet	144	square inches
square feet	0.092903	square meters
square feet	929.0341	square centimeters
square feet	3.587×10^{-8}	square miles
square feet	1/9	square yards
square feet/cu ft	3.29	sq m/cu m
square foot-square foot (moment of area)	20,736	square inch-square inch
square inches	1.273×10^{6}	circular mils
square inches	6.4516258	square centimeters
square inches	6.944×10^{-3}	square feet
square inches	645.2	square millimeters
square inches	10^{6}	square mils
square inches	7.71605×10^{-4}	square yards
square inches-inches sqd.	41.62	sq cm-cm sqd
square inches-inches sqd.	4.823×10^{-5}	sq feet-feet sqd
square kilometers	247.1	acres
square kilometers	10^{10}	square centimeters
square kilometers	10.76×10^{6}	square feet
square kilometers	1.550×10^{9}	square inches
square kilometers	10^{6}	square meters
square kilometers	0.3861006	square miles (U.S.)
square kilometers	1.196×10^{6}	square yards
square links (Gunter's)	10^{-5}	acres (U.S.)
square links (Gunter's)	0.04047	square meters
square meters	2.471×10^{-4}	acres (U.S.)
square meters	10^{4}	square centimeters
square meters	10.76387	square feet (U.S.)
square meters	1550	square inches

Multiply	*by*	*to obtain*
square meters	3.8610×10^{-7}	square miles (statute)
square meters	10^{6}	square millimeters
square meters	1.196	square yards (U.S.)
square miles	640	acres
square miles	2.78784×10^{7}	square feet
square miles	2.590	sq km
square miles	2.5900×10^{6}	square meters
square miles	3.098×10^{6}	square yards
square millimeters	1.973×10^{3}	circular mils
square millimeters	0.01	square centimeters
square millimeters	1.076×10^{-5}	square feet
square millimeters	1.550×10^{-3}	square inches
square mils	1.273	circular mils
square mils	6.452×10^{-6}	square centimeters
square mils	10^{-6}	square inches
square rods	272.3	square feet
square yard	2.1×10^{-4}	acres
square yards	8361	square centimeters
square yards	9	square feet
square yards	1296	square inches
square yards	0.8361	square meters
square yards	3.228×10^{-7}	square miles
square yards	8.361×10^{5}	square millimeters
statamperes	3.33560×10^{-10}	amperes (abs)
statcoulombs	3.33560×10^{-10}	coulombs (abs)
statcoulombs/kilogram	1.0197×10^{-6}	statcoulombs/dyne
statfarads	1.11263×10^{-12}	farads (abs)
stathenries	8.98776×10^{11}	henries (abs)
statohms	8.98776×10^{11}	ohms (abs)
statvolts	299.796	volts (abs)
statvolts/inch	118.05	volts (abs)/centimeter
statwebers	2.99796×10^{10}	electromagnetic cgs units of magnetic flux
statwebers	1	electrostatic cgs units of magnetic flux
stilb	2919	footlambert
stilb	1	int. candle cm^{-2}
stilb	3.142	lambert
stoke (kinematic viscosity)	10^{-4}	$meter^2$/second
stones (British)	6.350	kilograms
stones (British)	14	pounds
temp. (degs. C.) + 273	1	abs. temp. (degs. K.)
temps (degs. C.) + 17.8	1.8	temp. (degs. Fahr.)
temps. (degs. F.) + 460	1	abs. temp. (degs. R.)
temps. (degs. F.) − 32	5/9	temp. (degs. Cent.)

Multiply	*by*	*to obtain*
toises (French)	6	paris feet (pieds)
tons (long)	5.734×10^5	drams (avdp)
tons (long)	2.613×10^5	drams (troy)
tons (long)	1.568×10^7	grains
tons (long)	1.016×10^6	grams
tons (long)	1016	kilograms
tons (long)	3.584×10^4	ounces (avdp)
tons (long)	3.267×10^4	ounces (troy)
tons (long)	2240	pounds (avdp)
tons (long)	2722.2	pounds (troy)
tons (long)	1.12	tons (short)
Tons (metric) (T)	1000	kilograms
Tons (metric) (T)	2204.6	pounds
Tons (metric) (T)	1.1025	tons (short)
tons (short)	5.120×10^5	drams (avdp)
tons (short)	2.334×10^5	drams (troy)
tons (short)	1.4×10^7	grains
tons (short)	9.072×10^5	grams
tons (short)	907.2	kilograms
tons (short)	32,000	ounces (avdp)
tons (short)	29,166.66	ounces (troy)
tons (short)	2000	pounds (avdp)
tons (short)	2.430.56	pounds (troy)
tons (short)	0.89287	tons (long)
tons (short)	0.9078	Tons (metric) (T)
tons (short)/sq ft	9765	kg/sq meter
tons (short)/sq ft	13.89	pounds/sq inch
tons (short)/sq in	1.406×10^6	kg/sq meter
tons (short)/sq in	2000	pounds/sq inch
tons/sq mile	3.125	pounds/acre
tons/sq mile	0.07174	pounds/1000 sq ft
tons/sq mile	0.3503	grams/sq meter
tons/sq mile	350.3	kilograms/sq km
tons/sq mile	350.3	milligrams/sq meter
tons/sq mile	0.03503	milligrams/sq cm
tons/sq mile	0.03254	grams/sq ft
tons of water/24 hours	83.333	pounds of water/hr
tons of water/24 hours	0.16643	gallons/min
tons of water/24 hours	1.3349	cu ft/hr
torr (mm Hg, 0°C)	133.322	newton/meter^2
townships (U.S.)	23040	acres
townships (U.S.)	36	square miles
tuns	252	gallons
volts (abs)	10^8	abvolts

Multiply	*by*	*to obtain*
volts (abs)	3.336×10^{-3}	statvolts
volts (international of 1948)	1.00033	volts (abs)
volt/inch	.39370	volt/cm
watts (abs)	3.41304	BTU (mean)/hour
watts (abs)	0.0569	BTU (mean)/min
watts (abs)	0.01433	calories, kilogram (mean)/minute
watts (abs)	10^7	ergs/second
watts (abs)	44.26	foot-pounds/minute
watts (abs)	0.7376	foot-pounds/second
watts (abs)	0.0013405	horsepower (electrical)
watts (abs)	1.360×10^{-3}	horsepower (metric)
watts (abs)	1	joules/sec
watts (abs)	0.10197	kilogram-meters/second
watts (abs)	10^{-3}	kilowatts
watt-hours	3.415	British Thermal Units
watt-hours	3.60×10^{10}	ergs
watt-hours	2655	foot-pounds
watt-hours	859.85	gram-calories
watt-hours	1.34×10^{-3}	horsepower-hours
watt-hours	3.6×10^3	joule
watt-hours	0.8605	kilogram-calories
watt-hours	367.1	kilogram-meters
watt-hours	10^{-3}	kilowatt-hours
watt (international)	1.0002	watt (absolute)
watt/(cm^2)(°C/cm)	693.6	BTU/(hr)(ft^2)(°F/in)
wave length of the red line of cadmium	6.43847×10^{-7}	meters
webers	10^3	electromagnetic cgs units
webers	3.336×10^{-3}	electrostatic cgs units
webers	10^5	kilolines
webers	10^8	lines
webers	10^8	maxwells
webers	3.336×10^{-3}	statwebers
webers/sq in	1.550×10^7	gausses
webers/sq in	10^8	lines/sq in
webers/sq in	0.1550	webers/sq cm
webers/sq in	1,550	webers/sq meter
webers/sq meter	10^4	gausses
webers/sq meter	6.452×10^4	lines/sq in
webers/sq meter	10^{-4}	webers/sq cm
webers/sq meter	6.452×10^{-4}	webers/sq in
weeks	168	hours
weeks	10,080	minutes

Multiply	*by*	*to obtain*
weeks	604,800	seconds
yards	91.44	centimeters
yards	3	feet
yards	36	inches
yards	9.144×10^{-4}	kilometers
yards	0.91440	meters
yards	4.934×10^{-4}	miles (naut.)
yards	5.682×10^{-4}	miles (stat.)
yards	914.4	millimeters
years (sidereal)	365.2564	days (mean solar)
years (sidereal)	366.2564	days (sidereal)
years (tropical, mean solar)	365.2422	days (mean solar)
years (common)	8760	hours
years (tropical, mean solar)	8765.8128	hours (mean solar)
years (leap)	366	days
years (leap)	8784	hours
years (tropical, mean solar)	3.155693×10^{7}	seconds (mean solar)
years (tropical, mean solar)	1.00273780	years (sidereal)

2. BASIC AND SUPPLEMENTARY UNITS

A *meter (m)* is 1,650,763.73 wavelengths in vacuo of the radiation corresponding to the transition between the energy levels $2p_{10}$ and $5d_5$ of the krypton 86 atom.

A *kilogram (kg)* is the mass of the international prototype in the custody of the Bureau International des Poids et Mesures at Sevres in France.

A *second (sec)* is the interval occupied by 9,192,631,770 cycles of the radiation corresponding to the transition of the cesium-133 atom when unperturbed by exterior fields.

An *ampere* is the constant current that if maintained in two parallel rectilinear conductors of infinite length of negligible circular cross section and placed at a distance of one meter apart in vacuo would produce between these conductors a force equal to 2×10^{-7} newton per meter length.

A *kelvin* ($^\circ K$) is the degree interval of the thermodynamic scale on which the temperature of the triple point of water is 273.16 degrees.

A *candle* is such that the luminance of a full radiator at the temperature of solidification of platinum is 60 units of luminous intensity per square centimeter.

A *mole (mol)* is the amount of substance which contains as many elementary units as there are atoms in 0.012 kg of carbon-12. The elementary unit must be specified and may be an atom, an ion, an electron, a photon, etc., or a given group of such entities.

A *radian* is the angle subtended at the center of a circle by an arc of the circle equal in length to the radius of the circle.

A *steradian* is the solid angle that, having its vertex at the center of a sphere, cuts off an area of the surface of the sphere equal to that of a square with sides of length equal to the radius of the sphere.

3. DERIVED UNITS AND QUANTITIES

The *liter* was defined in 1901 as the volume of 1 kilogram of pure water at normal atmospheric pressure and maximum density equal therefore to 1.000028 dm^3. This 1901 definition applied for the purpose of the 1963 Weights and Measures Acts.

By a resolution of the 12th Conference General des Poids et Mesures (CGPM) in 1964 the word *liter* is now recognized as a special name for the dm^3, but is not used to express high precision measurements. It is used widely in engineering and the retail business, where the discrepancy of 28 parts in 1 million is of negligible significance.

A *newton (N)* is the force that, when applied to a body of mass of one kilogram, gives it an acceleration of one meter per second per second.

Stress is defined as the resultant internal force per unit area resisting change in the shape or size of a body acted on by external forces, and is therefore measured in *newtons per square meter* (N/m^2).

A *bar* is a pressure equivalent to 100,000 newtons acting on an area of one square metor.

A *joule (J)* is the work done when the point of application of a force of one newton is displaced through a distance of one meter in the direction of the force.

A *watt* is equal to one joule per second.

Dynamic viscosity is the property of a fluid whereby it tends to resist relative motion within itself. It is the shear stress, i.e., the tangential force on unit area, between two infinite horizontal planes at unit distance apart, one of which is fixed while the other moves with unit velocity. In other words, it is the shear stress divided by the velocity gradient, i.e., $(N/m^2) \div (m/sec/m) = N\ sec/m^2$.

Kinematic viscosity is the dynamic viscosity of a fluid divided by its density, i.e., $(N\ sec/m^2)/(kg/m^3) = m^2/sec$.

Density of heat flow rate (or heat flux) is the heat flow rate (W) per unit area, i.e., W/m^2.

Coefficient of heat transfer is the heat flow rate (W) per unit area per unit temperature difference, i.e., $W/m^{2\circ}C$.

Thermal conductivity is the quantity of heat that will be conducted in unit time through unit area of a slab of material of unit thickness with a unit difference of temperature between the faces; in other words, the heat flow rate (W) per unit area per unit temperature gradient, i.e., $W/[m^2(^\circ C/m)] = W/m^\circ C$.

The *heat capacity* of a substance is the quantity of heat gained or lost by the substance per unit temperature change, i.e., $J/^\circ C$.

Specific heat capacity is the heat capacity per unit mass of the substance, i.e., $J/kg^\circ C$.

Internal energy is the kinetic energy possessed by the molecules of a substance due to temperature and is measured in joules (J).

Specific internal energy (u) is the internal energy per unit mass of the substance, i.e., J/kg. When a small amount of heat is added at constant volume the increase in specific internal

energy is given by: $\mathrm{du} = c_\mathrm{v}\, dT$, where c_v is the specific heat capacity at constant volume, and dT is the increase in absolute temperature.

Specific enthalpy (h) is defined by the equation: $h = u + pv$, where p is the pressure and v is the specific volume. Specific enthalpy is measured in J/kg. When a small amount of heat is added to a substance at constant pressure, the increase in specific enthalpy is given by: $-dh = cp\, dT$, where cp is the specific heat capacity at constant pressure.

The *specific latent heat* of a substance is the heat gained per unit mass without an accompanying rise in temperature during a change of state at constant pressure. It is measured in J/kg.

The *entropy* (S) of a substance is such that when a small amount of heat is added, the increase in entropy is equal to the quantity of heat added (dQ) divided by the absolute temperature (T) at which the heat is absorbed; i.e., $dS = dQ/T$, measured in J/°K.

The *specific entropy* (s) of a substance is the entropy per unit mass, i.e., J/kg°K.

A *volt* is the difference of electric potential between two points of a conductor carrying a constant current of one ampere when the power dissipated is one watt.

A *weber* (Wb) is the magnetic flux through a conductor with a resistance of one ohm when reversal of the direction of the magnetic flux causes the transfer of one coulomb in the conductor loop.

Tesla: The magnetic flux density is the normal magnetic flux per unit area and is measured in *teslas*.

A *lumen*, the unit of luminous flux, is the flux emitted within unit solid angle of one steradian by a point source having a uniform intensity of one candle.

A *lux* is an illumination of one lumen per square meter.

Luminance is the luminous intensity per unit area of a source of light or of an illumination. It is measured in candles per square meter.

4. PHYSICAL CONSTANTS

Standard temperature and pressure (S.T.P.)	$= 273.15°\text{K}$ and $1.013 \times 10^5\ \text{N/m}^2$ $= 0°\text{C}$ and $1.013\ \text{bar}$ $= 0°\text{C}$ and $760\ \text{mm Hg}$
Molecular volume of ideal gas at S.T.P.	$= 22.41\ \text{liters/mol}$
Gas constant (R)	$= 8.314\ \text{J/mol°K}$
$^{RT}(273.15°\text{K})$	$= 2.271 \times 10^3\ \text{J/mol}$
Avogadro constant	$= 6.023 \times 10^{23}/\text{mol}$
Boltzmann constant	$= 1.3805 \times 10^{-23}\ \text{J/K}$
Faraday constant	$= 9.6487 \times 10^4\ °\text{C/mol}\ (= \text{A s/mol})$
Planck constant	$= 6.626 \times 10^{-34}\ \text{J sec}$
Stefan-Boltzman constant	$= 5.6697 \times 10^{-8}\ \text{W/m}^2\ \text{K}^4$
Ice point of water	$= 273.15°\text{K}\ (0°\text{C})$
Triple point of water	$= 273.16°\text{K}\ (0.01°\text{C})$
Speed of light	$= 2.998 \times 10^8\ \text{m/sec}$
Acceleration of gravity (standard) (Greenwich)	$= 9.80665\ \text{m/s}^2$ [take g as] $= 9.81188\ \text{m/s}^2$ [$9.81\ \text{m/s}^2$]
Universal constant of gravitation	$= 6.670 \times 10^{-11}\ \text{Newton m}^2/\text{kg}^2$
Mass of hydrogen atom	$= 1.6734 \times 10^{-27}\ \text{kg}$

5. PROPERTIES OF WATER

Temperature (°F)	Specific weight, γ (lb/ft^3)	Mass density, ρ (lb-sec^2/ft^4)	Dynamic viscosity, $\mu \times 10^5$ (lb-sec/ft^2)	Kinematic viscosity, $\nu \times 10^5$ (ft^2/sec)	Surface energy, $\sigma \times 10^3$ (lb/ft)	Vapor pressure, ρ (lb/in.^{2})	Bulk modulus, $E \times 10^{-3}$ (lb/in.^{2})
32	62.42	1.940	3.746	1.931	5.18	0.09	290
40	62.43	1.938	3.229	1.664	5.14	0.12	295
50	62.41	1.936	2.735	1.410	5.09	0.18	300
60	62.37	1.934	2.359	1.217	5.04	0.26	312
70	62.30	1.931	2.050	1.059	5.00	0.36	320
80	62.22	1.927	1.799	0.930	4.92	0.51	323
90	62.11	1.923	1.595	0.826	4.86	0.70	326
100	62.00	1.918	1.424	0.739	4.80	0.95	329
110	61.86	1.913	1.284	0.667	4.73	1.24	331
120	61.71	1.908	1.168	0.609	4.65	1.69	333
130	61.55	1.902	1.069	0.558	4.60	2.22	332
140	61.38	1.896	0.981	0.514	4.54	2.89	330
150	61.20	1.890	0.905	0.476	4.47	3.72	328
160	61.00	1.896	0.838	0.442	4.41	4.74	326
170	60.80	1.890	0.780	0.413	4.33	5.99	322
180	60.58	1.883	0.726	0.385	4.26	7.51	318
190	60.36	1.876	0.678	0.362	4.19	9.34	313
200	60.12	1.868	0.637	0.341	4.12	11.52	308
212	59.83	1.860	0.593	0.319	4.04	14.7	300

6. PERIODIC TABLE OF THE ELEMENTS (COMPLIMENTS OF THE LENOX INSTITUTE OF WATER TECHNOLOGY)

Groups ⇨ / Periods & sub-shells ⇩	I IA	2 IIA	3 IIIB	4 IVB	5 VB	6 VIB	7 VIIB	8	9 VIII	10	11 IB	12 IIB	13 IIIA	14 VIA	15 VA	16 VIA	17 VIIA	18 O
1 IS	1 H 1.00794 Hydrogen																1 H 1.00794 Hydrogen	2 He 4.00260 Helium
2 2s2p	3 Li 6.941 Lithium	4 Be 9.01218 Beryllium											5 B 10.811 Boron	6 C 12.011 Boron	7 N 14.0067 Nitrogen	8 O 15.9994 Oxygen	9 F 18.9984 Fluorine	10 Ne 20.179 Neon
3 3s3p	11 Na 22.9897 Sodium	12 Mg 24.305 Magnesium											13 AI 26.9815 Aluminum	14 Si 28.0855 Silicon	15 P 30.9738 Phosphorus	16 S 32.066 Sulfur	17 CI 35.4527 Chlorine	18 Ar 39.948 Argon
4 4s3d4p	19 K 39.098 Potassium	20 Ca 40.078 Calcium	21 Sc 44.9559 Scandium	22 Ti 47.88 Titanium	23 V 50.9415 Vanadium	24 Cr 51.996 Chromium	25 Mn 54.938 Manganese	26 Fe 55.847 Iron	27 Co 58.933 Cobalt	28 Ni 58.69 Nickel	29 Cu 63.546 Copper	30 Zn 65.39 Zinc	31 Ga 69,723 Gallium	32 Ge 69.561 Germanium	33 As 74.9216 Arsenic	34 Se 78.96 Selenium	35 Br 79.904 Bromine	36 Kr 83.80 Krypton
5 5s4d5p	37 Rb 85.468 Rubidium	38 Sr 87.62 Strontium	39 Y 88.9059 Yttrium	40 Zr 91.224 Zirconium	41 Nb 92.9064 Niobium	42 Mo 95.94 Molybdenum	43 Tc (98) Technetium	44 Ru 101.07 Ruthenium	45 Rh 102.906 Rhodium	46 Pd 106.42 Palladium	47 Ag 107.868 Silver	48 Cd 112.411 Cadmium	49 In 114.82 Indium	50 Sn 118.710 Tin	51 Sb 121.75 Antimony	52 Te 127.60 Tellurium	53 I 126.90 Iodine	54 Xe 131.29 Xenon
6 6s4f5d6p	55 Cs 132.905 Cesium	56 Ba 137.327 Barium	57 La 138.906 Lanthanum	72 Hf 178.49 Hafnium	73 Ta 180.948 Tantalum	74 W 183.85 Tungsten	75 Re 186.207 Rhenium	76 Os 190.2 Osmium	77 Ir 192.22 Iridium	78 Pt 195.08 Platinum	79 Au 196.97 Gold	80 Hg 200.59 Mercury	81 TI 204.383 Thallium	82 Pb 207.2 Lead	83 Bi 208.98 Bismuth	84 Po (209) Polonium	85 At (210) Astatine	86 Rn (222) Radon
7 7s5f6d	87 Fr (223) Francium	88 Ra (226) Radium	89 Ac (227) Actinium	104 Rf (261) Ruther-fordium	105 Ha (262) Dubnium	106 Sg (263) Seaborgium	107 Ns (262) Bohrium	108 Hs (265) Hassium	109 Mt (266) Meitnerium	110		112						

6 4f	58 Ce 140.116 Cerium	59 Pr 140.91 Praseody-mium	60 Nd 144.24 Neodymium	61 Pm (145) Promethium	62 Sm 150.35 Samarium	63 Eu 107.26 Europium	64 Gd 157.25 Gadolinium	65 Tb 158.925 Terbium	66 Dy 162.50 Dysprosium	67 Ho 104.930 Holmium	68 Er 167.26 Erbium	69 Tm 168.934 Thulium	70 Yb 173.04 Ytterbium	71 Lu 174.967 Lutetium
7 5f	90 Th 232.038 Thorium	91 Pa (231) Protactinium	92 U 238.029 Uranium	93 Np (237) Neptunium	94 Pu (244) Plutonium	95 Am (243) Americium	96 Cm (247) Curium	97 Bk (247) Berkelium	98 Cf (251) Californium	99 Es (252) Einsteinium	100 Fm (257) Fermium	101 Md (258) Mendelevium	102 No (259) Nobelium	103 Lr (262) Lawrencium

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