HANDBOOK OF ENVIRONMENTAL ENGINEERING VOLUME 6

Biosolids Treatment Processes

Edited by

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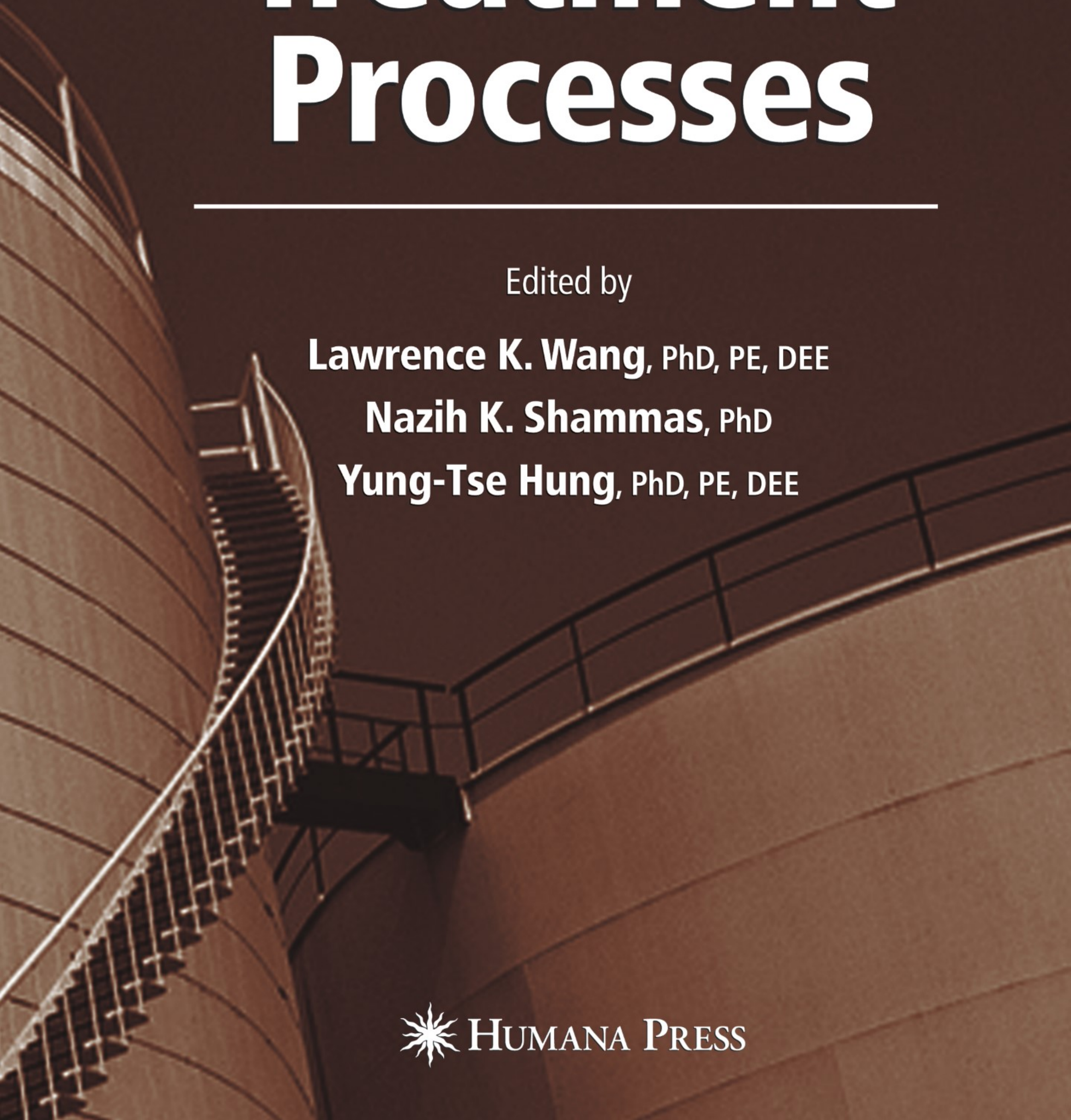

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Preface

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. Because 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 above three questions.

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.

Because 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 the 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. Conversion Factors for Environmental Engineers are attached as an appendix in this handbook for the convenience of international readers. 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 6, *Biosolids Treatment Processes*, is a sister book to Volume 7, *Biosolids Engineering and Management*. Both biosolids books have been designed to serve as basic biosolids treatment textbooks as well as comprehensive reference books. We hope and expect they will prove of equal high value to advanced undergraduate and graduate students, to designers of wastewater, biosolids, and sludge treatment systems, and to scientists and researchers. The editors welcome comments from readers in all of these categories. It is our hope that both books will not only provide information on the physical, chemical and biological treatment technologies, but will also serve as a basis for advanced study or specialized investigation of the theory and practice of individual biosolids management systems.

This book, Volume 6, *Biosolids Treatment Processes*, covers the topics of biosolids characteristics and quantity, gravity thickening, flotation thickening, centrifugation, anaerobic digestion, aerobic digestion, lime stabilization, low temperature thermal processes, high temperature thermal processes, chemical conditioning, stabilization, elutriation, polymer conditioning, drying, belt filter, composting, vertical shaft digestion, flotation, biofiltration, pressurized ozonation, evaporation, pressure filtration, vacuum filtration, anaerobic lagoons, vermicomposting, irradiation, and land application.

The sister book, Volume 7, *Biosolids Engineering and Management*, covers additional topics on sludge and biosolids transport, pumping and storage, sludge conversion to biosolids, waste chlorination for stabilization regulatory requirements, cost estimation, beneficial utilization, agricultural land application, biosolids landfill engineering, ocean disposal technology assessment, combustion and incineration, and process selection for biosolids management systems.

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.

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Characteristics and Quantity of Biosolids

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

This first chapter principally discusses the quantities and characteristics of biosolids produced by various wastewater treatment processes. There are several sources of wastewater biosolids; these biosolids can vary widely in characteristics and quantity. From the standpoint of quantity per unit of flow, the principal variables are the strength of the wastewater, whether chemicals are utilized in the process, and the degree of treatment.

The typical wastewater biosolids are classified as primary, biological, and chemical. The biological biosolids are waste-activated sludges (WAS) and fixed film biosolids from various types of filters and rotating biological contactors. The activated sludge might have primary biosolids incorporated into the biomass when primary clarifiers are not used. Chemical residues might be produced simultaneously with primary biosolids or biological biosolids through the addition of metal salts for precipitation of phosphorus, or they can be made in a separate tertiary treatment stage. Lime is sometimes used in the primary treatment stage and also in the tertiary stage, when softening of the effluent is required for reuse. The reader is referred to US Environmental Protection Agency (US EPA) Manual for Phosphorus Removal (1) for a discussion of the production and dewatering characteristics of chemical biosolids.

In some cases, well designed biosolids handling systems were actually marginal in operation because of inaccurate estimates of wastewater treatment loadings of 5-d biochemical oxygen demand (BOD_5), and total suspended solids (TSS), and the subsequent biosolids production. These problems may occur for the following reasons:

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Table 1
US Municipal Biosolids Production by POTW Size

POTW size (MGD)	No. of POTWs	Sludge produced (dry t/yr)	Total (%)
<2.5	14,168	1,189,810	17
2.5–5	631	515,504	8
5–10	352	588,445	9
10–20	187	622,478	9
20–50	125	924,896	13
50–100	40	676,091	10
>100	41	2,324,274	34

Source: US EPA.
ton × 0.9072 = tonne (t = 0.9072T).
MGD × 0.0438 = m^3/s.

a. Low estimate of unit biosolids yield/unit of chemical oxygen demand (COD) or BOD_5 removal.
b. Use of average weekly or monthly BOD_5 and TSS inputs.
c. No allowance for the normal peak day/average discharge characteristics of larger industrial facilities.
d. Inaccurate estimate of primary treatment efficiency.
e. Effects of BOD_5 and TSS recycle ignored or underestimated.
f. Seasonal discharges of BOD_5 and TSS overlooked.

The quantity of biosolids produced in US municipal wastewater treatment plants was estimated by the US EPA based on its 1982 survey. Table 1 presents these data for all sizes of treatment plants. The number of Publicly Owned Treatment Works (POTWs) in a particular size category is also given.

It is interesting to note that the smallest plants, <2.5 MGD (0.11 m^3/s), represent 91% of the POTWs and produce less than 17% of the biosolids. In contrast, the largest plants, >100 MGD (4.38 m^3/s) represent less than 0.3% of the facilities and produce more than 34% of the biosolids. A representative survey of United States facilities was performed in 1980 by the US EPA's Office of Solid Waste to determine the choice of biosolids use/disposal options by plant size. The results are shown in Table 2. The "Other" category frequently means a lagoon or temporary storage facility. Note that small to medium sized facilities frequently select some form of land use/disposal option compared with the large sized facilities, which more frequently use incineration.

When either evaluating or selecting a biosolids treatment process, one must keep in mind the inherent influence of the earlier wastewater and biosolids treatment processes as well as the subsequent use or disposal practices. Choice of a utilization/disposal process is in turn strongly influenced by local, state, and federal regulations.

Any particular process cannot be evaluated without considering the other processes involved in the overall wastewater/solids handling system. This evaluation or selection can be a complex procedure because of the large number of possible combinations of unit processes available for wastewater treatment and biosolids thickening, stabilization, conditioning, dewatering, and ultimate use/disposal. Figure 1 shows the unit processes most commonly used to perform most of these functions. An evaluation

Table 2
Percent Distribution of Biosolids Disposal/Utilization Practices for US POTWs

Practice	Small POTWs (<1 MGD)	Medium POTWs (1–10 MGD)	Large POTWs (>10 MGD)	Total of all POTWs
Landfill	31	34	12	15
Incineration	1	1	32	27
Land application	39	38	21	24
Distribution and marketing	11	17	19	18
Ocean disposal	1	–	4	4
Other	17	10	12	12
Total (%)	100	100	100	100

Source: US EPA.
MGD × 0.0438 = m^3/s.

procedure should start at the bottom of the figure with the use/disposal options and work backwards to come to a decision on the particular process.

Before selecting a biosolids management option, the design engineer should consult all regulations for a particular use/disposal system to observe, if a minimum requirement is mandatory. For example, federal regulations state specific mandatory requirements for stabilization, pathogenic content, and concentration of heavy metals (2,3). Other states minimum level requirements usually apply which can vary widely from state to state.

The characteristics and quantity of generated biosolids are, in turn, affected not only by the raw wastewater quality, but also by the management options chosen for the treatment of the wastewater streams. Biosolids quantity and quality are covered in this first chapter of the book. The various biosolids treatment and utilization/disposal options are discussed in the following chapters.

2. PRIMARY BIOSOLIDS

Most wastewater treatment plants use primary sedimentation to remove readily settleable solids from raw wastewater. In a typical plant with primary sedimentation and a conventional-activated sludge process for secondary treatment, the dry weight of primary solids is roughly 50% of that for the total solids. For several reasons, primary biosolids are usually easier to manage than biological and chemical biosolids. First, primary biosolids are readily thickened by gravity, either within a primary sedimentation tank or within a separate gravity thickener. In comparison with biological and many chemical biosolids, primary biosolids with low conditioning requirements can be mechanically dewatered rapidly. Further, the dewatering device will produce a drier cake and give better solids capture than it would for most biological and chemical sludges (4).

2.1. Estimation of Primary Biosolids Production

Primary biosolids production is typically within the range of 800–2500 lb/MG (100–300 mg/L) of wastewater. A basic approach to estimating primary biosolids production for a particular plant is by computing the quantity of TSS entering the primary sedimentation tank and assuming an efficiency of removal. When site-specific data are

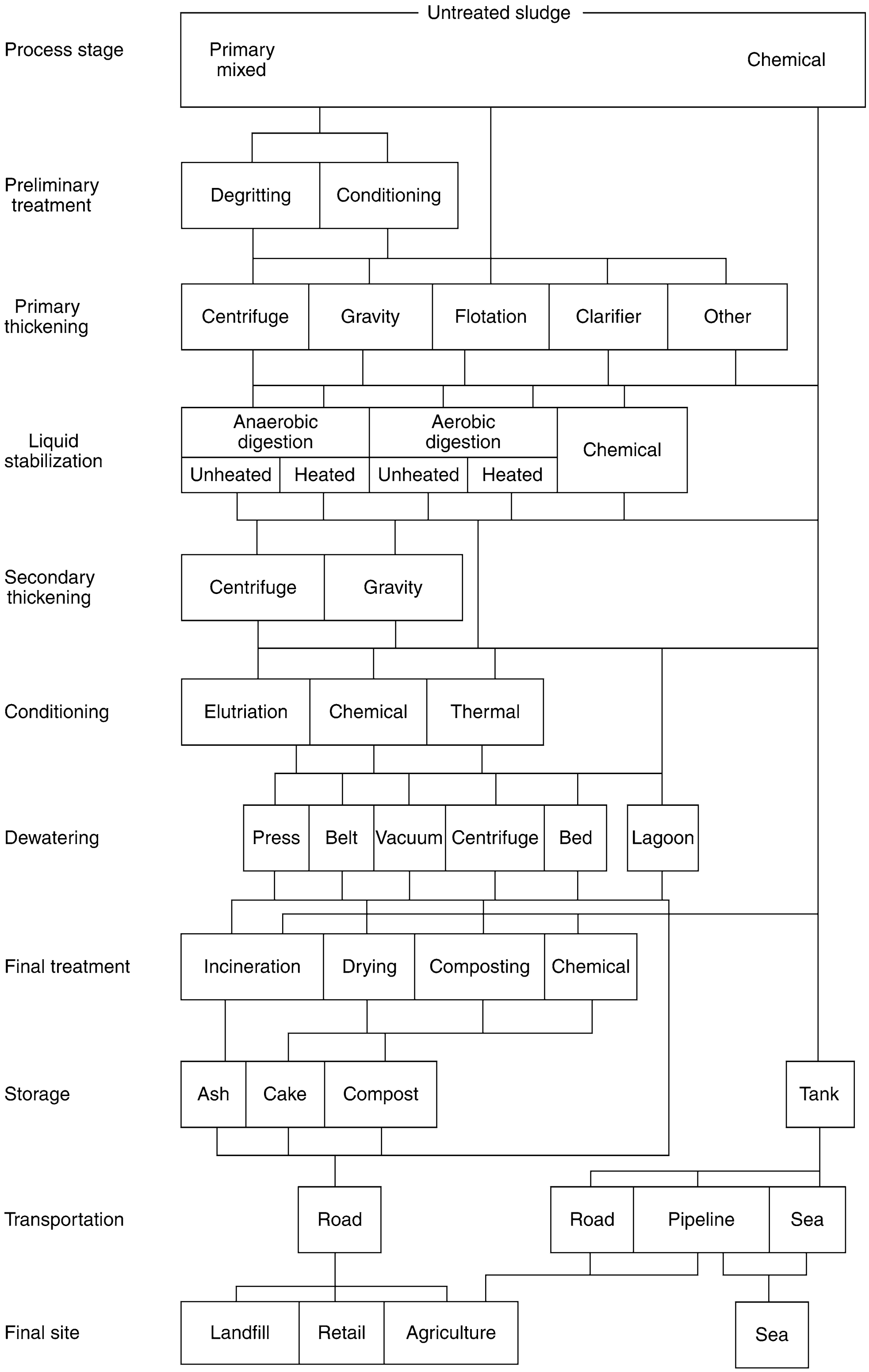

Fig. 1. Biosolids management and processing options (*Source:* US EPA).

not available for influent TSS, estimates of 0.15–0.24 lb/capita/d (0.07–0.11 kg/capita/d) are commonly used (5).

Usually, removal efficiency of TSS in the primary sedimentation tank is in the 50–65% range (6). An efficiency of 60% is frequently used for estimating purposes, subject to the following conditions:

a. The biosolids are produced in treatment of a domestic wastewater without major industrial loads.
b. The sludge contains no chemical coagulants or flocculants.
c. No other biosolids, for example, trickling filter biosolids, have been added to the influent wastewater.
d. The biosolids contain no major sidestreams from biosolids processing.

As an example, if a designer estimates the TSS entering the primary clarifier as 0.2 lb/capita/d (0.09 kg/capita/d), and the removal efficiency of the clarifier as 60%, the estimated primary biosolids production is 0.12 lb/capita/d (0.054 kg/capita/d). The raw primary biosolids (RPB) production can be easily determined from the total flow and the influent and effluent TSS of the primary clarifier. Care should be taken to ensure that the influent sample, which does not contain recycled solids, is the same as the primary clarifier influent. Some adjustment of the influent is necessary to account for recycled solids removed by primary clarification. Even with good operation, recycled TSS can amount to 15–20% of influent TSS, and the BOD_5 recycle is usually 8–15% of the influent BOD_5.

The influent loadings and resulting biosolids production should be analyzed and developed into a frequency plot, which would indicate the frequency of a specific TSS and BOD_5 influent loading (kg/d) vs time frequency (%). Similar graphs should be plotted for RPB produced and primary effluent (PE) BOD_5 (kg/d vs time frequency [%]). A mass balance of the overall process should be prepared to ensure accounting of all TSS and BOD_5.

Unlike secondary biosolids, the volatility of primary biosolids might vary considerably from day-to-day and seasonally. This is particularly true of wastewater systems with combined sewers and/or substantial infiltration and inflow. The domestic and commercial discharges of volatile suspended solids (VSS) would not vary widely throughout the year. Some short-term increases may be noted as a result of "first-flush" effects during sudden wet weather conditions. First-flush effects occur with the transport of accumulated solids in the sewers and street washing where there are combined sewers. Where there are seasonal or variable industrial discharges, the VSS might vary widely, and the primary removals of TSS and BOD_5 may also vary, depending on the nature of the solids.

Designers should anticipate that reductions in the primary biosolids volatile content generally, will be accompanied by proportionally more biosolids, eventhough there might be only a small or no increase in the VSS loading. An example follows:

a. **Dry weather.** TSS = 100,000 lb/d, VSS = 78,000 lb/d = 78%.
b. **Wet weather.** TSS = 130,000 lb/d, VSS = 78,000 lb/d = 60%.

The biosolids production at the lower volatile content is 30% higher than the dry weather biosolids quantity. Although the lower volatile content biosolids is somewhat easier to handle because of the grit content, the biosolids handling design should anticipate the higher quantity of solids. In existing plants, the past operating records should

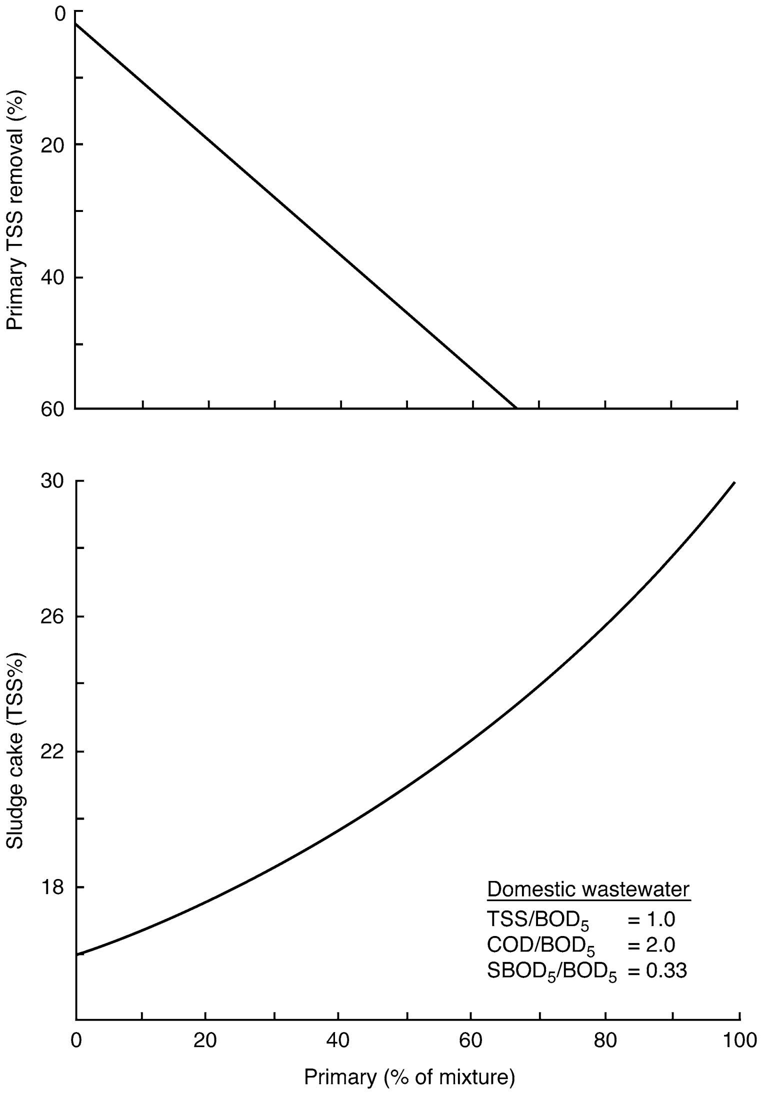

Fig. 2. Primary TSS Removal and dewatered cake concentration for various Primary and WAS Mixtures (*Source:* US EPA).

be scrutinized to determine if there is a significant variation in the actual biosolids volatility and the biosolids quantity projected.

The efficiency of primary clarification is important because not only are the primary biosolids easier to handle, but the unit yield (kg/kg) of secondary biosolids is partially dependent on the TSS/ BOD_5 in the clarified mixture. Figure 2 presents the yield and the dewatered cake concentration of various RPB:WAS ratios. When primary clarifiers are not used, the total quantity of solids produced is lower than RPB + WAS, but the water retention characteristics of the biological solids increase. Although the absolute values shown in Fig. 2 vary, depending on biosolids characteristics and the mechanical equipment used, the general relationship holds.

In cases where the highest possible cake solids are required, good primary treatment should be provided. The primary clarifier requirements can be experimentally determined using laboratory settling tests, if the wastewater is available or by evaluating the performance characteristics of existing units at various flow rates. The clarifier performance is strongly influenced by the surface overflow rate (SOR [$m^3/m^2/d$ or $gal/ft^2/d$]) and the clarifier sidewater depth (SWD). Good performance of circular primary clarifiers will be achieved when:

a. English units.

$\text{SOR, max} \leq 10\ \text{SWD}^2$ (SWD = 6 ≤ 10 ft).

$\text{SOR, max} \leq 100\ \text{SWD}$ (SWD = 10 ≤ 15 ft).

b. Metric units.

$\text{SOR, max} \leq 4.5\ \text{SWD}^2$ (SWD = 2 ≤ 3 m).

$\text{SOR, max} \leq 12.25\ \text{SWD}$ (SWD = 3 ≤ 5 m).

For rectangular clarifiers, the length of the flow path is the most important to overcome inlet disturbances. The depth of the basin is also significant. For basins less than 30 m long, the length to width ratio is ≥5:1. Basins that are 30–65 m long and 4–5 m deep will provide excellent results, even at rates up to and exceeding 67.2 $m^3/m^2/d$ (1650 $gal/ft^2/d$).

2.2. *Factors Affecting Solids Removal*

Suspended solids removal efficiency in primary sedimentation depends to a large extent on the nature of the solids. It is difficult to generalize about the effect that industrial suspended solids can have removal efficiency, but an example illustrates that the effect can sometimes be dramatic (4). At North Kansas City, Missouri, a municipal plant serves residential customers and numerous major industries, including food processing, paint manufacturing, soft-drink bottling, paper manufacturing, and grain storage and milling. Raw wastewater entering the plant had a 15-d average suspended solids concentration of 1140 mg/L that was attributable to the industries. Primary sedimentation removed 90% of these solids. The quantity of primary biosolids was, therefore, about 8000 lb/MG (1000 mg/L) of wastewater treated. This value is several times the normal one for domestic wastewater. On two of the 15 d, removal exceeded 14,000 lb/MG (1700 mg/L) (7).

Home garbage grinders can significantly increase the suspended solids load on a wastewater treatment plant. These solids are largely settleable. Estimates of the increased primary biosolids resulting from the use of garbage grinders range from 25 to more than 50% (5,8,9).

Operating experience shows clearly that the amount of biosolids withdrawn from the primary sedimentation tank is highly increased, when biosolids treatment process sidestreams such as digester supernatant, elutriate, and filtrates or centrates, and other biosolids like WAS are recycled to the primary sedimentation tank. Quantifying the solids entering and leaving the primary clarifier by all streams is an important tool for estimating primary biosolids production, when recycled biosolids and biosolids process sidestreams contribute large quantities of solids.

Table 3
Quantities of Solids Removed in a Primary Sedimentation Tank

Sludge type	No chemical addition[b]	Chemical addition[a] Lime[c]	 Alum[d]	 Iron[e]
Suspended solids (lb/MG)	1041	1562	1562	1562
Chemical solids (lb/MG)	–	2082	362	462
Total sludge production (lb/MG) [kg/m^3]	1041 [0.13]	3644 [0.44]	1924 [0.23]	2024 [0.24]

Source: US EPA.
[a] Assumes 10 mg/L influent phosphorus concentration (as P) with 80% removed by chemical precipitation.
[b] Assumes 50% removal of 250 mg/L influent TSS in primary sedimentation.
[c] 125 mg/L $Ca(OH)_2$ added to raise pH to 9.5.
[d] 154 mg/L $A1_2(SO_4)_3 \cdot 14\ H_2O$ added.
[e] 84 mg/L $FeCl_3$ added.

When chemicals are added to the raw wastewater for removal of phosphorus or coagulation of nonsettleable solids, large quantities of chemical precipitates are formed. The quantity of chemical solids produced in chemical treatment of wastewater depends upon the type and amount of chemical(s) added, chemical constituents in the wastewater, and performance of the coagulation and clarification processes. It is difficult to predict accurately the quantity of chemical solids that will be produced. Classical jar tests are favored as a means for estimating chemical biosolids quantities. The quantities of suspended solids and chemical solids removed in a primary sedimentation tank (10,11), that is processing wastewater which has been treated by lime, aluminum sulfate, or ferric chloride addition are shown in Table 3.

Peak rates of primary biosolids production can be several times the average. Peak solids production levels also vary from one plant to another. Three studies of primary biosolids production rates are summarized and presented here. At Ames, Iowa, (12) the wastewater is basically of domestic origin. A university contributes about 30% of the volumetric and mass loads. Storm runoff is collected and kept separate from the domestic wastewater. For 21 yr of record, the suspended solids load in the peak month of each year were divided by the yearly average. The average of these ratios was 1.37. The average for comparison of peak days and peak months more than 10 yr of record was 1.59. Thus, in a typical year, the maximum daily flow would be about 1.37 × 1.59, or 2.2 times the average. Therefore, the maximum day's biosolids production was expected to follow a similar pattern and was estimated to be 2.2 times the average value.

A study used data from Chicago, Cleveland, Columbus, Syracuse, Rochester, and several other large American cities (13) to show a typical relationship between the peak raw municipal wastewater solids loads entering a plant and duration of time that these peaks persist. This relationship is shown graphically in Fig. 3. The curve is appropriate for large cities with a number of combined sewers on flat grades. The peaks occur at least partly because solids deposited in the sewers at low flows are flushed out by storm flows.

Data were collected more than a 5-yr period from the West Point plant at Seattle, Washington and used in the third study (14). Peak primary biosolids load of 4–10-d durations were compared with average loads. The duration of 4 d was selected because

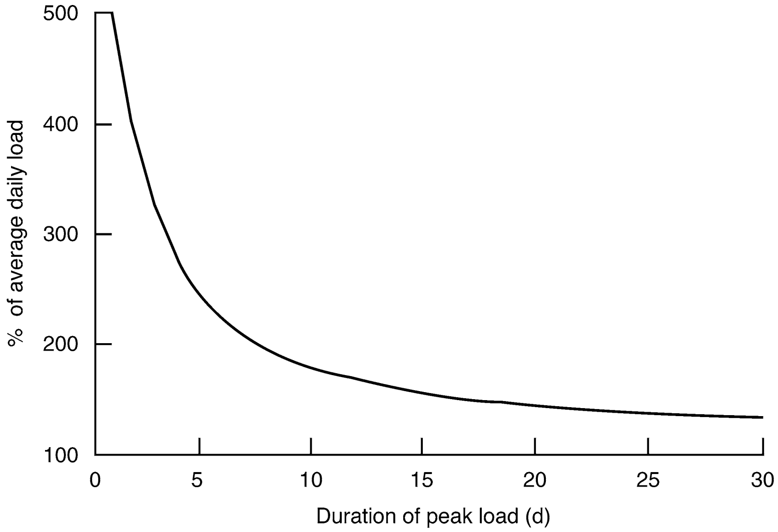

Fig. 3. Relationship between peak solids loading and duration of peak for some large US cities (*Source:* US PEA).

it appeared to be highly significant to digester operations at this plant, and because loads tended to drop after about 4 d of heavy loading. The highest 4-d primary biosolids production was four times more than the normal production from the plant's service area. Main contributors to the peak load were:

a. **Solids deposits in the sewers.** These deposits were resuspended during high flows and carried to the treatment plant. The computer-operated storage system, which minimizes combined sewer overflows, apparently contributed to solids deposition/reentrainment.
b. **Storm inflow.** Measurements of TSS in storm drainage fluctuate widely but often show over 200 mg/L suspended solids. A large portion of the west point service area contains combined sewers.
c. **Biosolids conditioning and dewatering.** Problems in these processes have caused the sidestreams to contain more solids than usual.

2.3. Composition and Characteristics of Primary Biosolids

Most primary biosolids can be concentrated readily within the primary sedimentation tanks. Several authors state that a 5–6% solids concentration is attainable when biosolids are pumped from well-designed primary sedimentation tanks (6,13,16,17). However, values both higher and lower than the 5–6% range are common. Conditions that influence primary biosolids concentration include:

a. If wastewater is not degritted before it enters the sedimentation tanks, the grit might be removed by passing the RPBs through cyclonic separators. However, these separators do not function properly with biosolids concentrations more than 1% (18).
b. If the biosolids contain large amount of fine nonvolatile solids, such as silt, from storm inflow, a concentration of well over 6% might sometimes be attained (14,19).
c. Industrial loads might strongly affect primary biosolids concentration. For example, at a plant, receiving soil discharged from a tomato canning operation, primary biosolids with a

17% solids concentration, of which 40% is volatile, was recorded. Normal primary biosolids at this plant had a solids concentration of 5–6% solids (60–70% volatile) (20).

d. Primary biosolids might float when buoyed up by gas bubbles generated under anaerobic conditions. Conditions favoring gas formation include: warm temperatures; solids deposits within sewers; strong septic wastes; long detention times for wastewater solids in the sedimentation tanks; lack of adequate prechlorination; and recirculating sludge liquors (21). To prevent the septic conditions that favor gas formation, it might be necessary to strictly limit the storage time of sludge in the sedimentation tanks. This is done by increasing the frequency and rate of primary sludge pumping (22).
e. If biological sludges are mixed with the wastewater, generally, a lower primary sludge concentration will result.

Table 4 lists a number of primary sludge characteristics. In many cases, ranges and/or "typical" values are given. In the absence of recirculating biosolids process sidestreams, the percent of volatile solids in the primary Biosolids should approximate the percent volatile suspended solids in the influent wastewater. A volatile solids content lower than 70% usually indicates the presence of storm water inflow, biosolids processing sidestreams, a large amount of grit, residues from a water filtration plant that was discharged to the sanitary sewer, low volatile solids from industrial waste, or wastewater solids that have a long detention time in the sewers.

Primary biosolids always contain some grit, even when the wastewater has been processed through degritting. Where screenings are comminuted and returned to the wastewater flow, the fragmented screenings appear in the primary biosolids. Smaller plastic and rubber items that pass through screens also appear in the primary biosolids. Primary biosolids typically contain more than 100 different anaerobic and facultative species of bacteria (23). Sulfate-reducing and oxidizing bacteria, worm and fly eggs, and pathogenic microorganisms are typically present.

3. BIOLOGICAL BIOSOLIDS

3.1. General Characteristics

Biological biosolids are produced by treatment processes such as activated sludge, trickling filters, and rotating biological contactors. Quantities and characteristics of biological biosolids vary with the metabolic and growth rates of the various microorganisms present in the biosolids. The quantity and quality of biosolids produced by the biological process are intermediate between that produced in no-primary systems and that produced in full-primary systems, in cases when fine screens or primary sedimentation tanks with high overflow rates are used. Biological biosolids containing debris such as grit, plastics, paper, and fibers, will be produced at plants lacking primary treatment. Normally, plants with primary sedimentation produce a fairly pure biological biosolids. The concentrations and, therefore, the volumes of waste biological biosolids are highly affected by the method of operation of the clarifiers. Generally, biological biosolids are more difficult to thicken and dewater than primary biosolids and most chemical biosolids.

3.2. Activated Sludge

3.2.1. Activated Sludge Production-Dry Weight Basis

Activated sludge has numerous variations such as extended aeration, oxidation ditch, pure oxygen, mechanical aeration, diffused aeration, plug flow, contact stabilization,

Table 4
Primary Biosolids Characteristics

Characteristic	Range of values	Typical value	Comments
pH	5–8	6	–
Volatile acids (mg/L as acetic acid)	200–2000	500	–
Heating value (Btu/lb)[a]	6800–10,000	–	Depends upon volatile content and sludge decomposition, reported values are on a dry weight basis
	–	10,285	Sludge 74% volatile
	–	7600	Sludge 65% volatile
Specific gravity of individual solid particles	–	1.4	Increases with increased grit, silt, and so on
Bulk specific gravity (wet)	–	1.02	Increases with sludge thickness and with specific gravity of solids
		1.07	Strong sewage from a system of combined storm and sanitary sewers
BOD_5/VSS ratio	0.5–1.1	–	–
COD/VSS ratio	1.2–1.6	–	–
Organic N/VSS ratio	0.05–0.06	–	–
Volatile content (percent/weight of dry solids)	64–93	77	Value obtained with no sludge recycle, good degritting; 42 samples, standard deviation 5
	60–80	65	
	–	40	Low value caused by severe storm inflow
	–	40	Low value caused by industrial waste
Cellulose (percent/weight of dry solids)	8–15	10	–
	–	3.8	–
Hemicellulose (percent/weight of dry solids)	–	3.2	–
Lignin (percent/weight of dry solids)	–	5.8	–
Grease and fat (percent/weight of dry solids)	6–30	–	Ether soluble
	7–35	–	Ether extract
Protein (percent/weight of dry solids)	20–30	25	–
	22–28	–	
Nitrogen (percent/weight of dry solids)	1.5–4	2.5	Expressed as N

(Continued)

Table 4 ***(Continued)***

Characteristic	Range of values	Typical value	Comments
Phosphorus (percent/weight of dry solids)	0.8–2.8	1.6	Expressed as P_2O_5. Divide values as P_2O_5 by 2.29 to obtain values as P
Potash (percent/ weight of dry solids)	0–1	0.4	Expressed as K_2O. Divide values by 1.2 to obtain values as K

Source: US EPA.

[a] 1 Btu/lb = 2.32 kJ/kg.

complete mix, step feed, nitrifying activated sludge, and so on (6,24–29). The quantity of WAS is affected by two parameters: the dry weight of the biosolids and the concentration of the biosolids. This section describes how the dry weight of activated sludge production may be predicted.

The most important variables in predicting WAS production are the amounts of organics removed in the process, the mass of microorganisms in the system, the biologically inert suspended solids in the influent to the biological process, and the loss of suspended solids to the effluent. These variables can be assembled into two simple and useful equations:

$$P_X = (Y)(S_r) - (k_d)(M) \tag{1}$$

$$\mathrm{WAS}_T = P_X + I_{\mathrm{NV}} - E_T \tag{2}$$

where P_X is the net growth of biological solids expressed as volatile suspended solids (VSS) (lb/d or kg/d); Y is the gross yield coefficient (lb/lb or kg/kg); S_r is the substrate (e.g., BOD_5) removed (lb/d or kg/d); k_d is the decay coefficient (d^{-1}); M is the system inventory of microbial solids (VSS) (lb or kg); WAS_T is the waste-activated sludge production (lb/d or kg/d); I_{NV} is the nonvolatile suspended solids fed to the process (lb/d or kg/d); E_T is the effluent suspended solids (lb/d or kg/d).

These equations, as stated or with slight variations, have been widely used. Equation (1) dates back to 1951 (25). However, different terms and symbols have been used by various authors in expressing Eqs. (1) and (2). Table 5 summarizes some of the terminology that has evolved. The technical literature reflects some inconsistency in terminology with the term "M." Test results reported by various authors and presented in Table 5 were derived on the basis of "M" defined as mixed liquor VSS only.

To use Eq. (1), it is necessary to obtain values of Y and k_d. Whereas Table 6 summarizes several reported values for these parameters, it is best to determine Y and k_d on an individual waste stream whenever possible.

To use Eq. (2), it is necessary to estimate I_{NV} nonvolatile influent solids, and E_T, effluent suspended solids. Generally, the following are included within the term I_{NV}:

a. Nonvolatile solids in influent sewage, including recycle biosolids liquors.
b. Chemical precipitates—for example, aluminum phosphates- when alum is added to the activated sludge process.
c. Stormwater solids that are not removed in previous processes (31).

Table 5
Alternate Names and Symbols for Equations (1)

Symbol	Name	Dimensions	Other symbols for similar quantities	Other common names for similar quantities
P_X	Biological solids production	Mass/time	ΔX_V, dX/dT, A, S, dM/dT, R_g	Accumulation, net growth, excess microorganisms production
Y	Gross yield coefficient[a]	Mass/mass	a, Ks, C	Yield coefficient, synthesis coefficient, growth-yield coefficient
s_r	Substrate removal	Mass/time	dF/dT, S, B, F_i, R	Food, utilization, load
k_d	Decay constant	1/time	b, K_d, K_e	Endogenous respiration, maintenance energy, auto-oxidation
M	Microbial solids inventory	Mass	S, X, X_V	Microbial mass, solids under aeration, solids inventory, mixed liquor solids

Source: US EPA.

[a] The letter Y has also been used for the net yield coefficient P_x/s_r. The net yield coefficient is quite different from the gross yield coefficient.

d. Normal nonvolatile content of the activated sludge. In the absence of biosolids liquors, chemical precipitates, and stormwater, activated sludge will be about 80% volatile (less in extended aeration) at most municipal treatment plants.

To compute E_T, a small value such as 10 mg/L TSS should be used.

3.2.2. Factors That Influence the Production Of WAS

1. **Effect of sludge age and F/M ratio.** Equation (1) can be rearranged to show the effect of the sludge age (θ_m):

$$P_X = (Y)\ (S_r)/1 + (k_d)\ (\theta_m) \tag{3}$$

where θ_m is the M/P_X = Sludge age (d). Similarly, Eq. (1) can be rearranged to show the effect of the food-to-microorganism ratio (F/M):

$$P_X = (Y)\ (S_r) - (k_d)\ (S_r)/(C_2)\ (F/M) \tag{4}$$

where C_2 is the coefficient to match units of S_r and "F" in F/M; if S_r is BOD_5 removed (influent – effluent), then C_2 is BOD_5 removal efficiency, about 0.9; F/M is the food-to-microorganism ratio = BOD_5 applied daily/VSS (mass) in system.

As θ_m increases and F/M decreases, the biological solids production P_X decreases. Biosolids handling is expensive, and costs can be reduced by using high values of θ_m or low values of F/M. However, there are offsetting cost factors, such as increases in the aeration tank volume needed, oxygen requirements for the aerobic biological system, and so on. As seasons change, so may the optimum θ_m and F/M also change to correspond to maximum wastewater treatment efficiency. Therefore, it is desirable to be able

Table 6
Values of Yield and Decay Coefficients for Computing WAS

Cross yield coefficient[a]	Decay coefficient[b]	Type of wastewater	Scale of plant	Aeration	Temperature (°C)	Sludge age (d)[c]	BOD_5 removal calculation
0.5	0.055	Primary effluent	Bench	Air	19–22	2.8–22	Influent
0.70	0.04	Primary effluent	Pilot	Oxygen	Not stated	1–4	Influent – effluent
0.67	0.06	Primary effluent	Full	Air	18–27	1.2–8	Influent – effluent
0.73	0.075	Primary effluent	Pilot	Air	10–16	1–12	Influent – effluent
0.94	0.14	Primary effluent (wastewater includes dewatering liquors)	Pilot	Air	15–20	0.5–8	Influent – soluble effluent
0.73	0.06	Primary effluent	Pilot	Oxygen	18–22	2.5–17	Influent – effluent
0.5	Not calculated (negligible)	Primary effluent (military base)	Pilot	Air	0–7	Long[d]	Influent
0.74	0.04	Primary effluent (much industry)	Pilot	Oxygen	17–25	2.1–5	Influent – soluble effluent
1.57	0.07	Raw degritted including dewatering liquors	Pilot	Air	15–20	0.6–3	Influent – soluble effluent
1.825	0.20	Raw degritted	Bench	Air	4–20	1–3	Soluble influent – soluble effluent
0.65	0.043	Raw degritted	Bench	Air	20–21	11 and up[d]	Influent – effluent
0.70	0.048	Raw degritted	Bench	Air	20–21	Long[d]	Influent – effluent
0.54	0.014	Raw degritted	Full	Air	Not stated	Long[d]	Influent – effluent
1.1	0.09	Raw	Full	Air	Not stated	1.1–2.4	Influent – effluent

Source: US EPA.

Note: All values in this table are for an equation of the type $P_x = Ys_r - k_dM$.

[a] Gross yield coefficient Y, lb (kg) VSS/lb (kg) BOD_5.

[b] Decay coefficient k_d (d^{-1}).

[c] Mean cell residence time or sludge age θ_m, measured as mass of mixed liquor VSS divided by biological solids production P_x. Note that coefficients might be somewhat different if total system inventory of VSS (mixed liquor VSS + VSS in clarifiers) is used rather than just mixed liquor VSS.

[d] Extended aeration.

to operate across a range of conditions. Obviously, trial-and-error calculations are required to determine the least costly system.

2. **Effect of nitrification.** Nitrification is the bio-oxidation of ammonia nitrogen and organic nitrogen to the nitrite and nitrate forms (32). Compared with processes that are designed for carbonaceous (BOD_5, COD) oxidation only, stable nitrification processes operate at long sludge ages (θ_m) and low food-to-microorganism ratios (F/M). Also, nitrification processes are often preceded by other processes that remove much of the BOD_5 and suspended solids (SS) (26). As a result, activated sludge in a nitrification mode generally, produces less WAS than conventional activated sludge processes. However, there is an additional component to nitrification biosolids, the net yield of nitrifying bacteria (Y_N). This might be estimated at 0.15 lb SS/lb of total Kjeldahl nitrogen (organic + ammonia) removed (33). Y_N varies with temperature, pH, dissolved oxygen (DO), and cell residence time. However, detailed measurements of Y_N are not ordinarily required for biosolids facility design because the yield of nitrifying bacteria is small (34). For example, if Y_N is 0.15 and if the nitrifying process removes an ammonia nitrogen concentration of 20 mg/L and an organic nitrogen concentration of 10 mg/L then nitrification would add $0.15 \times (20 + 10) = 4.5$ mg of nitrifying bacteria/L of wastewater (38 lb/MG). These quantities are small compared with other biosolids. In single-stage nitrification processes, the biosolids production figures must also include the solids produced from the carbonaceous oxidation, computed at the θ_m and F/M of the nitrifying system.
3. **Effect of feed composition.** The type of wastewater that is fed to the activated sludge process has a major influence on the gross yield (Y) and decay (k_d) coefficients. Many industrial wastes contain large amount of soluble BOD_5 but small amount of suspended or colloidal solids. Normally, these wastes have lower Y coefficients than are obtained with domestic PE. On the other hand, wastes with large amounts of solids, relative to BOD_5, either have higher Y coefficients or require adjustments to reflect the influent inert solids. Even among soluble wastes, different compositions will cause different yields.
4. **Effect of** DO **concentration.** Various DO levels have been maintained in investigations of activated sludge processes. Very low DO concentrations, for example, 0.5 mg/L, in conventional activated sludge systems do appear to cause increased solids production, even when other factors are held constant (35). However, there is vigorous disagreement concerning solids production at higher DO levels. Some investigators state that use of pure oxygen instead of air reduces sludge production. This is attributed to the high DO levels attained through the use of pure oxygen (36,37). Other investigators in well-controlled investigations have concluded that if at least 2 mg/L DO is maintained in air-activated sludge systems, then air and oxygen systems produce the same yield at equivalent conditions (such as food-to-microorganism ratio) (38,39).
5. **Effect of temperature.** The coefficients Y (gross yield) and k_d (decay) are related to biological activity and, therefore, might vary because of temperature of the wastewater (32,34). This variation has not been well documented in pilot studies and process investigations. One study obtained no significant difference because of temperature over the range 39–68°F (4–20°C) (40). However, others have observed significant differences within the same temperature range. Sometimes a simple exponential (Arrhenius) equation is used for temperature corrections to Y and k_d. For instance, it has been stated that chemical and biochemical rates double with an 18°F (10°C) rise in temperature. Exponential equations have been found to be accurate for pure cultures of bacteria, but are quite inaccurate when applied to Y and k_d for the mixed cultures found in real-activated sludge (41,42).

   For the design engineer, the following guidelines are recommended:

   a. Wastewater temperatures in the range of 59–72°F (15–22°C) might be considered to be a base case. Most of the available data are from this range. Within this range, there is no need to make temperature corrections. Any variations in process coefficients across this temperature range are likely to be small in comparison to uncertainties caused by other factors.

b. If wastewater temperatures are in the range of 50–59°F (10–15°C), the same k_d value as for 59–75°F (15–22°C) should be used, but the Y-value should be increased by 26%. This is based on experiments that compared systems at 52°F (11°C) and 70°F (21°C). In these tests, k_d was the same, but *Y* was 26% higher. On a COD basis, Y was found to be 0.48 at 38°F (11°C) and 0.38 at 56°F (21°C) (43).
c. If wastewater temperatures are less than 50°F (10°C), increased biosolids production should be expected (44), but the amount of increase cannot be accurately predicted from available data. Under such conditions, there is a need for pilot-scale process investigations.
d. If wastewater temperatures are more than 72°F (22°C), values of the process coefficients from the range 59–72°F (15–22°C) might be used for design. The resulting design may be somewhat conservative.

6. **Effect of feed pattern.** Various feed patterns for the activated sludge process include contact stabilization, step feeding, conventional plug-flow, and complete-mix. For design purposes, it appears to be best to ignore the feed pattern when estimating solids production.

3.2.3. Peak Rate of WAS Production

Peak solids production occurs because of unfavorable combinations of the elements in Eqs. (1), (3), and (4), presented previously:

$$P_X = (Y)\,(S_r) - (k_d)\,(M) \tag{1}$$

$$P_X = (Y)\,(S_r)/1 + (k_d)\,(\theta_m) \tag{3}$$

$$P_X = (Y)\,(S_r) - (k_d)\,(S_r)/(C_2)\,(F/M) \tag{4}$$

All of these equations predict that solids production (P_X) increases with increases in S_r and *F*/*M* and decreases with increases in the mass of organisms and θ_m. Also P_X increases if the gross yield coefficient (*Y*) increases or if the decay coefficient (k_d) decreases. Each of these factors that tend to increase P_X will occur, within limits, in practice. To compute peak solids production, the following conditions should be assumed:

a. Peak substrate removal (S_r). If high efficiency of biological wastewater treatment is maintained at peak pollutant loading, then S_r represents organics removal at maximum load. If S_r is computed on a BOD_5 removal basis, then the maximum BOD_5 removal should be used. The duration of peak solids production will match the duration of the peak load.
b. Minimum value of θ_m or maximum *F*/*M*. This allows the operator to select θ_m or *F*/*M* to obtain the best possible effluent. The design average condition might be *F*/*M* = 0.3, but an operator might obtain better results at *F*/*M* = 0.5 for some specific conditions at a particular treatment plant.
c. Maximum likely value of *Y*.
d. Minimum likely value of k_d.

Also, a temperature allowance should be made if wastewater temperatures less than 59°F (15°C) might occur during peak loads. Solids inventory reductions are an additional type of non-steady state condition that the designer should anticipate. It is occasionally necessary for treatment plant operators to reduce the mass of microorganisms

(M) in the liquid treatment process by wasting activated sludge. Wasting activated sludge helps the operator to maintain a constant F/M in the face of reduced BOD_5 loadings. The wastewater-BOD_5 load can drop rapidly if a treatment plant serves vacation areas or industries. Wasting activated sludge also allows the operator to take aeration tanks, clarifiers, and so on, out of service to limit solids on clarifiers, and to prevent major loss of solids to the effluent and to inhibit the growth of undesirable microorganisms such as scum-causing actinomycetes (45). Further, by reducing M, the operator can more readily optimize bioflocculation, thereby, minimizing effluent solids, and can control air or oxygen requirements.

To accomplish the desired inventory reduction, solids handling facilities must have the capacity to accept the wasted solids. For wastewater treatment plants without major known BOD_5 and SS loading variations, allowance should be made in designing solids processing facilities for the wasting of an additional 2% of M/d and lasting up to 2 wk. Such plants include those serving stable domestic populations. Industrial loads would be either small or unusually stable.

For plants with major seasonal variations in loads, allowance should be made for wasting an additional 5% of M/d and lasting for up to 2 wk. Such plants serve resort areas, college towns, and so on. A similar allowance should be made for plants that practise nitrification during only part of the year. Lastly, for plants with major weekday-to-weekend variations of more than 2:1 in BOD_5 load, and medium or high food-to-microorganism ratios of more than 0.3 during the high loads, allowance should be made for a 1-d biosolids wasting of up to 25% of M. The plant should also be able to handle wasting of 5% of M/d and lasting for 2 wk. Plants in this category serve major industrial systems, large office complexes, schools, and ski areas.

Because inventory reductions are not generally practised during peak loading periods, these previously-discussed capacity allowances should be added to average solids production. The maximum rate of WAS production is determined by whichever is higher: production during peak loading or the sum of average production plus inventory reduction allowances.

Occasionally, biosolids are wasted in a pattern so that M increases at some times and decreases at others. An example of such a pattern is the withdrawal of WAS only during the daytime. The Tapia, California, Water Reclamation Plant uses this pattern to obtain good process control (46). Uses of such patterns will, of course, increase the maximum rate at which WAS must be removed.

3.2.4. Measurements of Biosolids Yield Coefficients

Pilot studies and full-scale operating records can provide better data for establishing biosolids production design criteria than any general compilation of data from other locations. Measurements of the biosolids yield coefficients are of two basic types. First, both the gross yield Y and the decay k_d might be determined. Second, observed net yields alone might be used.

Equations (1), (3), and (4) are used when the food-to-microorganism ratio F/M and the sludge age, θ_m might be expected to vary in the prototype plant. To use these equations, it is necessary to determine the two biosolids yield coefficients, Y and k_d. To establish

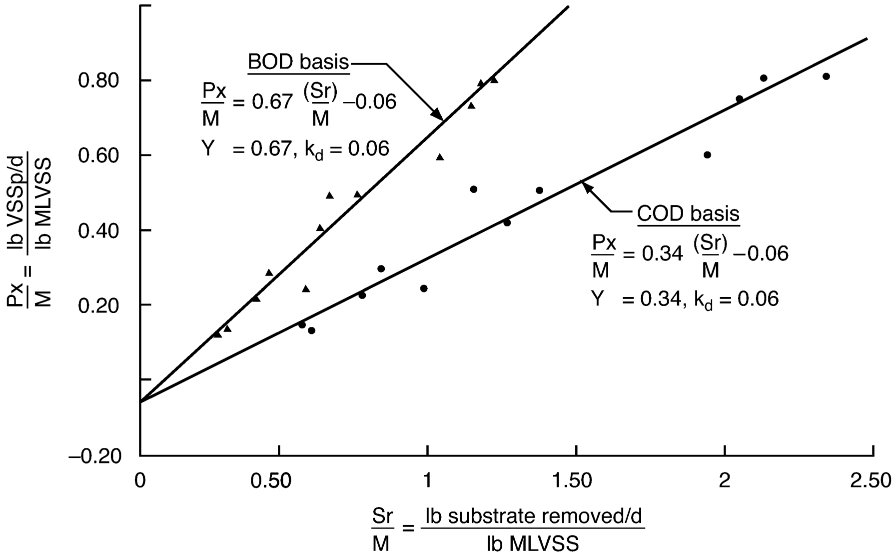

Fig. 4. Net growth rate curve (*Source:* US EPA).

these two coefficients, solids production must be measured under at least two different conditions of *F*/*M* and θ_m. Equation (1) can be rearranged slightly to Eq. (5):

$$\frac{P_X}{M} = Y\ (S_r/M)\ - k_d \tag{5}$$

where P_X/M is the net growth rate = $1/\theta_m$/d; S_r/M is the lb (kg) BOD_5 removed per day/lb (kg) VSS.

This equation provides a basic straight-line relationship between P_X/M and S_r/M. For each condition of operation, P_X/M and S_r/M are calculated and plotted, and a straight line is drawn through the points. As shown in Fig. 4, the slope of the line is the yield coefficient (*Y*), and the intercept represents the decay coefficient (k_d).

If the design conditions of S_r/M or θ_m are known and if solids production can be measured under these conditions, then it is not necessary to determine both *Y* and k_d. Instead, a simple observed net yield might be calculated. Equations (1) and (3) are easily rearranged to show:

$$Y_{\text{obs}} = \frac{P_X}{S_r} = Y - k_d/(Sr/M) = \frac{Y}{1+(k_d)\,(\theta_m)} \tag{6}$$

where Y_{obs} is the net yield coefficient = lb(kg) VSS produced/lb(kg) substrate (e.g., BOD_5) removed.

Net yield coefficients are often reported in the literature. They are directly applicable only under the conditions of S_r/M and θ_m that occurred during the experiments; they are meaningless unless S_r/M or θ_m are measured also. For gathering data from pilot plants or existing plants for use in establishing biosolids yield coefficients, several precautions should be exercised. Either automatic DO control should be used in the test or ample air or oxygen should be provided to ensure that the mixed liquor DO concentration is more than 2 mg/L at all times. Data from widely differing temperatures should not be plotted on the same graph to determine *Y* and k_d. Instead, data from each temperature range should be used to determine Y and k_d for that range. Each condition of S_r/M or θ_m should be maintained long enough to obtain stable operation. To assure system stability, a period of time equal to three times the sludge age should elapse between tests. The designer should use the term I_{NV} in Eq. (2) to correct the effect of sidestreams. The percent volatile content of the solids produced should be recorded. This will be useful in computing the total solids in the biosolids.

3.2.5. Concentration of WAS

The volume of biosolids produced by the process is directly proportional to the dry weight and inversely proportional to the thickness or solids concentration in the waste biosolids stream. Values for WAS concentration can vary, in practice, across a range from 1000 to 30,000 mg/L SS (0.1–3% SS).

An important variable that can affect WAS concentration is the method of biosolids wasting. A number of different methods are illustrated in Fig. 5. Sludge solids might be wasted from the clarifier underflow. It has been argued that wasting solids from the mixed liquor should improve control of the process (6,47). In this case, waste sludge is removed from the activated sludge process at the same concentration as the mixed liquor suspended solids, about 0.1–0.4%. This low concentration can be a disadvantage because a large volume of mixed liquor must be removed to obtain a given wastage on a dry weight basis. The most common arrangement involves sludge wasting from the clarifier underflow, because the concentration of sludge there is higher than in the mixed liquor. Subsequent discussions in this section are based on sludge wasting from the clarifier underflow.

The two primary factors that affect WAS concentration are the settleability of the biosolids and the solids loading rate to the sedimentation tank. These two factors have been considered in detail in the development of solids flux procedures for predicting the clarifier underflow concentration of activated sludge (48).

Various factors that affect biosolids settleability and the clarifier biosolids loading rate include:

a. **Biological characteristics of the biosolids.** These characteristics might be partially controlled by maintenance of a particular mean sludge age or *F/M*. High concentrations of filamentous organisms can sometimes occur in activated sludge. Reduction of these organisms through sludge age or *F/M* control helps to produce more concentrated clarifier underflow.

b. **Temperature.** As wastewater temperatures are reduced, the maximum attainable clarifier underflow biosolids concentration (C_u) is also reduced as a result of increased water density. Also, temperature can affect the setting properties of the biosolids.

c. **Solids flux.** The solids flux is the solids load from the mixed liquor divided by the clarifier area (e.g., lb/d/ft^2). Higher rates of solids flux require that clarifiers be operated at lower solids concentration.

d. **Limits of biosolids collection equipment.** Because of the pseudoplastic and viscous nature of WAS, some of the available biosolids collectors and pumps are not capable of smooth, reliable operation when C_u exceeds about 5000 mg/L.

e. **Heavy suspended solids in the biosolids.** If raw wastewater, instead of primary sedimentation tank effluent, is fed to the activated sludge process, usually higher C_u values result. Chemicals added to the wastewater for phosphorus and suspended solids removal might similarly affect C_u. However, such additional solids will also increase the solids load to the clarifiers.

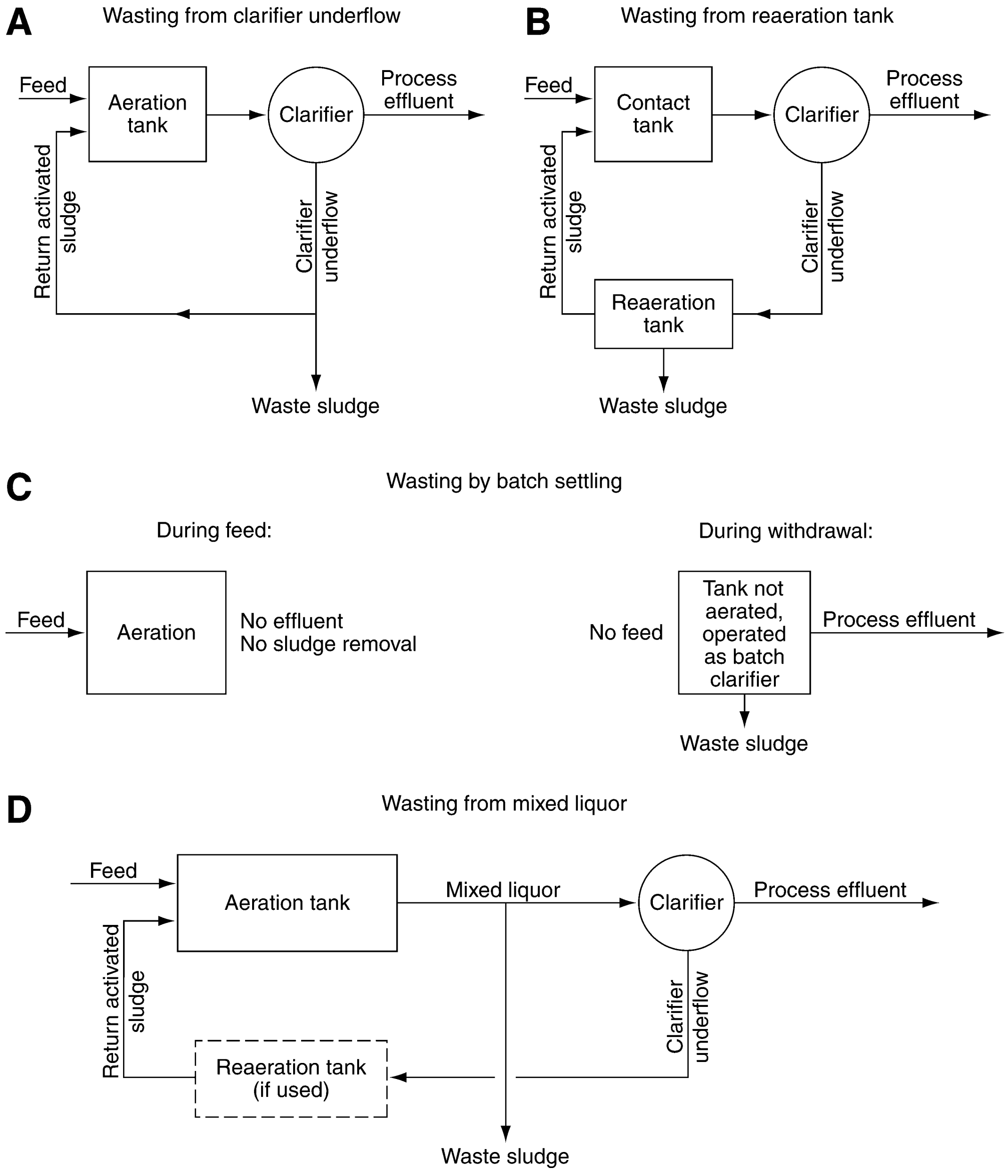

Fig. 5. Biosolids wasting methods (*Source:* US EPA).

3.2.6. Other Properties of Activated Sludge

Table 7 contains several reported measurements of the composition and properties of activated sludge solids. Comparing Table 7 with that of Table 4 for primary biosolids, activated sludge contains higher amount of nitrogen, phosphorus, and protein; the grease, fats, and cellulose amount, and specific gravity are lower.

Several types of microorganisms are present in large numbers in activated sludge. Floc-forming (zoogleal) bacteria include species of *Zoogloea, Pseudomonas, Arthrobacter, and Alcaligenes*. Activated sludge also contains filamentous microorganisms such as *Sphaerotilus*, *Thiothrix*, *Bacillus*, and *Beggiatoa* (49). Various protozoa are present, including ciliates and flagellates.

Table 7
Activated Sludge Characteristics

Characteristic	Range of values	Typical value	Comments
pH	6.5–8	–	Can be less in high purity oxygen systems or if anaerobic decomposition begins
		5.5	Baltimore, MD
Heating value (Btu/lb) [kJ/kg]	–	6540 [15,200]	Increases with percent volatile content
Specific gravity of individual solid particles	–	1.08	Sludge 74% volatile
Bulk specific gravity	–	$1 + 7 \times 10^{-8} \times C$	C is suspended solids concentration in mg/L
Color	–	Brown	Some grayish sludge has been noted. Activated sludge becomes black upon anaerobic decomposition
COD/VSS ratio	–	2.17	–
Carbon/nitrogen ratio	–	12.9	Baltimore, MD
	–	6.6	Jasper, IN
		14.6	Richmond, IN
	–	5.7	Southwest plant, Chicago, IL
	–	3.5	Milwaukee, WI (heat dried)
Organic carbon (percent/weight of dry solids)	17–41	–	Zurich, Switzerland
	23–44	–	Four plants
Nitrogen [percent/weight of dry solids (expressed as N)]	4.7–6.7	–	Zurich, Switzerland
	–	5.6	Chicago, IL
	2.4–5	–	Four plants
	–	6	Milwaukee, WI
Phosphorus, percent/weight of dry solids as P_2O_5 (divide by 2.29 to obtain Phosphorus as P)	3–3.7	–	Zurich, Switzerland
	–	7	Chicago, IL
	2.8–11	–	Four plants
	–	4	Milwaukee, WI
Potassium, percent/weight of dry solids as K_2O (divide by 1.2 to obtain Potassium as K)	0.5–0.7	–	Zurich, Switzerland
	–	0.56	Chicago, IL

(Continued)

Table 7 ***(Continued)***

Characteristic	Range of values	Typical value	Comments
	–	0.41	Milwaukee, WI
Volatile solids, percent/weight of dry solids (ash % is 100 – volatile %)	61–75	–	Zurich, Switzerland
	–	63	
	62–75	–	
	59–70	–	Four plants
	–	76	Renton, WA (Seatle Metro) 1976 average
	–	88	San Ramon, CA (Valley community Services District), 1975 average
	–	81	Central plant, Sacramento county, CA July 1977–June 1978 average
Grease and fat (percent/weight of dry solids)	5–12	–	Ether extract
Cellulose (percent/ weight of dry solids)		7	Includes lignin
Protein (percent/ weight of dry solids)	32–41	–	–

Source: US EPA.

3.3. Trickling Filter Biosolids

Trickling filters are widely used in municipal wastewater treatment. This section covers trickling filters that are used with clarifiers. When a clarifier is not used, the trickling filter effluent is usually fed to an activated sludge process.

3.3.1. Trickling Filter Biosolids Production—Dry Weight Basis

Trickling filter microorganisms are biochemically similar to microorganisms that predominate in activated sludge systems. Consequently, solids production from trickling filters and from activated sludge systems is roughly similar when compared on the basis of pounds of solids produced per pound of substrate removed. However, there are differences between the two systems, with respect to solids production prediction methodology and the pattern of biosolids wasting. Attempts have been made to develop solids production models consistent with biological theory (50–52). However, empirical methods are usually used for design purposes. Table 8 presents biosolids yields observed at several treatment plants and from one long-term pilot study. These data are primarily based on heavily loaded filters.

Table 8
Trickling Filter Solids Production

	Unit solids production[a]							
Plant	Total BOD_5 basis[b]	IT-ES BOD_5 basis[c]	IT-ES COD basis[d]	SS basis[e]	VSS basis[f]	Solids (volatile %)	BOD_5 load[g]	Media
Stockton, California[h]								*Plastic (27 ft²/ft³)*
Average month	0.74	0.67	0.43	1	0.94	77	27	
Highest month	1.01	0.92	0.60	1.17	1.08	86	73	
Lowest month	0.49	0.48	0.30	0.61	0.60	64	15	
Sacremento, California[h]								*Plastic*
Nine noncanning months average	–	–	–	1.01	1	78	–	
Highest month	–	–	–	1.09	1.09	83	–	
Three canning months average	–	–	–	1.2	1.24	76	–	
Dallas, TX	0.42	–	–	–	–	–	–	Rock
Dallas, TX	0.65	–					–	Rock
Livermore, California	1.10[i]	–	–	1.39	1.51	84	57	Rock 2–4 in.
San Pablo, CA	–	–	–	1.39	–	–	199	Plastic (29 ft²/ft³)
Seatle, WA[j]	–	0.8–0.9	–	1	–	–	30–250	Plastic various

Source: US EPA.

[a] Solids production includes both waste sludge (clarifier underflow) and clarifier effluent solids.

[b] Pounds volatile suspended solids (VSS) per pound BOD_5 removed (same as kg/kg). BOD_5 removal based on total (suspended + dissolved) measurements.

[c] Pounds VSS per pound BOD_5 removed. BOD_5 removal based on influent total – effluent soluble (IT-ES) measurements.

[d] Pounds VSS per pound COD removed. COD removal based on influent total minus effluent soluble measurements.

[e] Pounds total SS produced per pounds SS applied.

[f] Pounds VSS produced per pound VSS applied.

[g] Pounds total BOD_5 applied/d/1000 ft^3 of media.

[h] Stockton and Sacramento plants have heavy industrial loads about August to October from fruit and vegetable canneries.

[i] Roughing filter. For BOD_5 basis, BOD_5 removal was computed by $BOD_{5,\ in}$ minus (0.5 times unsettled $BOD_{5,\ out}$). 1971 average data.

[j] Pilot studies. SS basis was found to describe data well over a wide range of loadings. Wastewater included some industrial load and recycle liquors from dewatering digested sludge.

Equations that relate the production of suspended material in a trickling filter can be developed in a form similar to that used in predicting activated sludge production. The main difference lies in the term used to define the quantity of microorganisms in the system. In long-term studies of trickling filter performance, Merrill (52) assumed that the total mass of microorganisms present in the system was proportional to the media surface area. The resulting equation for volatile solids production was:

$$P_X = Y'(S_r) - k'_d A_m \tag{7}$$

where P_X is the net growth of biological solids (VSS) (lb/d or kg/d); Y' is the gross yield coefficient (lb/lb or kg/kg); k'_d is the decay coefficient (d^{-1}); S_r is the substrate (e.g., BOD_5) removed, lb/d or kg/d = BOD_5 in minus soluble effluent BOD_5; A_m is the total media surface area in reactor (ft^2 or m^2).

The production of trickling filter biosolids requiring subsequent biosolids handling might be expressed:

$$\mathrm{WTFB} = P_X + \mathrm{I}_{\mathrm{NV}} - E_T \tag{8}$$

where WTFB is the waste trickling filter biosolids production (lb/d or kg/d); I_{NV} is the nonvolatile suspended solids fed to the process (lb/d or kg/d); E_T is the effluent suspended solids (lb/d or kg/d).

The coefficients Y' and k'_d for Eq. (7) are obtained for a particular system by computing the slope and intercept of a line of the best fit through plotted data points for P_X/A_m vs S_r/A_m. VSS production data for three different trickling filter media designs are given on Fig. 6.

Nitrification in trickling filters causes a synthesis of nitrifying bacteria. However, as in activated sludge, the quantity is small. A value of 25 lb/MG (3 mg/L) has been suggested for design purposes (53). This quantity must be added to the other solids produced by the trickling filter.

It is known that temperature and loading rate affect biosolids production. The quantities of excess biosolids produced in low-rate trickling filters are lower than those for high-rate filters or for the activated sludge process. The lower rate of solids accumulation may be attributable to the grazing activities of protozoa. The activity of the protozoa is reduced at lower temperatures (50).

Peak biosolids loads are produced by trickling filters. These may be because of variations in influent load, rapid climatic changes, and/or biochemical factors that cause unusually large amount of biomass to peel off from the media. The term "sloughing" is used by some authorities to include steady state as well as peak solids discharges. Others restrict the term "sloughing" to unusually large discharges. In any case, peak solids loads must be considered. In low-rate filters especially, there are seasonal variations in solids production. Slime accumulates in the trickling filter during winter operation and the filter unloads the slime in the spring when the activity of the microorganisms is increased (50).

The amount of solids requiring biosolids treatment depends on sedimentation performance, which is usually 50–90% removal of suspended solids. Sedimentation performance is improved by careful design, light loads, tube settlers, and coagulation and flocculation (22,52,54).

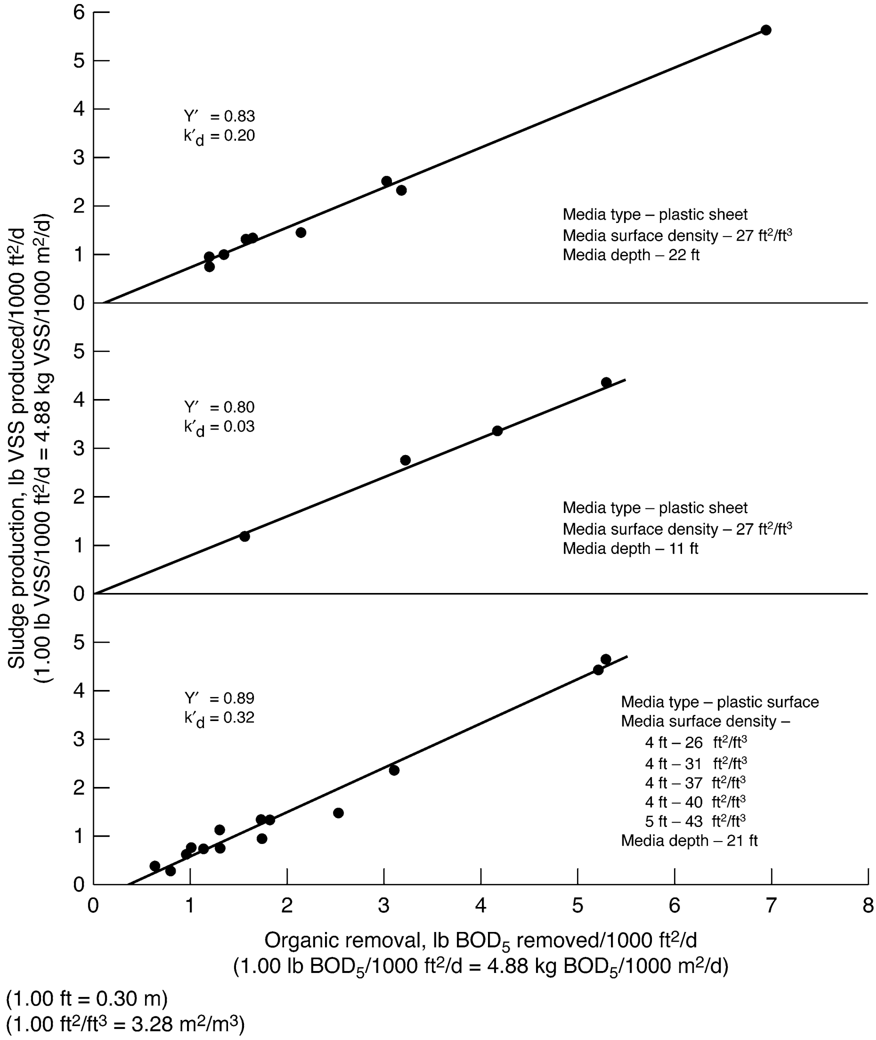

Fig. 6. VSS production data for three trickling media designs (*Source:* US EPA).

3.3.2. Concentration of Trickling Filter Biosolids

Trickling filter biosolids loadings on the secondary sedimentation tank are typically low – 5–10% of observed solids loads to activated sludge sedimentation tanks. Trickling filter biosolids also have better thickening properties than activated sludge. Consequently, trickling filter biosolids can be withdrawn at higher concentration than WAS. The solids flux method for predicting biosolids concentration might be used with trickling filter biosolids (48). This method requires measurement of initial solids settling velocity vs solids concentration. Such relationships have been reported for at least one trickling filter process (52).

3.3.3. Properties of Trickling Filter Biosolids

Table 9 contains a few analyses of trickling filter biosolids properties. The microbial population that inhabits a trickling filter is complex and includes many species of algae,

Table 9
Trickling Filter Biosolids Composition

Property	Value
Volatile content (percent of total solids)	64–86
Nitrogen (percent of total solids)	1.5–5
Phosphorus as P_2O_5 (percent of total solids)	2.8
	1.2
Fats (percent of total solids)	6
Grease (percent of total solids)	0.03
Specific gravity of individual solid particles	1.52
	1.33
Bulk specific gravity (wet)	1.02
Color	Grayish brown
	Black

Source: US EPA.

bacteria, fungi, protozoa, worms, snails, and insects. Filter flies and their larvae are often present in large numbers around trickling filters.

3.4. Biosolids from Rotating Biological Contactors

Rotating biological contactors (RBCs) are used for the same basic purposes as activated sludge and trickling filters: to remove BOD_5 and suspended solids and, where necessary, to nitrify. The RBC process uses a tank in which wastewater, typically PE, contacts plastic media in the shape of large discs. Bacteria grow on the discs. The discs rotate slowly on horizontal shafts; the bacteria are alternately submerged in the wastewater and exposed to air. Excess bacteria slough from the discs into the wastewater. After contacting the bacteria, the wastewater flows to a sedimentation tank, where the excess bacteria and other wastewater solids are removed. These removed solids are RBC biosolids. RBC biosolids are roughly similar in quantity by dry weight, nutrient content, and other characteristics, to trickling filter biosolids.

Published data is available on RBC biosolids production rate from pilot- and full-scale municipal installations (55–59). At Peewaukee, WI, TSS production has been reported to be 0.62–60.82 lb of TSS/lb BOD_5 (0.62–0.82 kg TSS/kg) removed. The final sedimentation tank removed 70–83% of these solids as biosolids. The biosolids had a concentration of 1.5–5% solids. Other investigations of municipal and industrial waste applications have concluded that biosolids production for the RBC process amounts to 0.4–0.5 lb of TSS/lb of BOD_5 (0.4–0.5 kg TSS/kg BOD_5) removed (60–62).

3.5. Coupled Attached-Suspended Growth Biosolids

There are several installations of coupled attached and suspended growth processes in the United States. Usually, these dual processes are installed where nitrification is required or where strong wastes must be treated. The attached growth reactor is a trickling filter or a rotating biological reactor. Its role is to reduce the load on the suspended growth process. The suspended growth process uses an aeration tank and a final clarifier. Usually, flow recirculation is practised around the attached growth reactor. Several reports describe these processes and note that the biosolids are similar to activated sludge, both in quantity and in characteristics (8,53,63–65).

3.6. Denitrification Biosolids

Denitrification is a biological process for the removal of nitrate from wastewater. An electron donor, carbon in PE or methanol, is added to the nitrate-bearing wastewater. Denitrifying bacteria extract energy for growth from the reaction of nitrate with the electron donor:

$$\text{Nitrate} + \text{electron donor (reduced state)} \rightarrow \text{nitrogen gas} \\ + \text{oxidized electron donor} + \text{energy}$$

Denitrification has been extensively studied, and denitrification processes have been built into municipal plants. Denitrifying bacteria can grow either in a suspended growth system similar to activated sludge or in an attached growth system similar to a trickling filter. Biosolids production for ordinary nitrified domestic waste is roughly 300 lb/MG (30 mg/L) of wastewater treated (33).

4. CHEMICAL BIOSOLIDS

When chemicals are added to the raw wastewater for removal of phosphorus or coagulation of nonsettleable solids, larger quantities of biosolids are formed (54,66,67). The quantity of solids produced in the chemical treatment of wastewater depends upon the type and amount of chemical(s) added, the chemical constituents in the wastewater, and the performance of the coagulation and clarification processes. It is difficult to predict accurately the quantity of chemical solids that will be produced. Jar tests are preferred as a means for estimating chemical biosolids quantities.

Table 10 provides estimated quantities of suspended and chemical solids removed in a hypothetical primary sedimentation tank processing wastewater that has been treated with lime, aluminum sulfate, or ferric chloride. The use of polyelectrolytes might highly enhance the solids capture in the clarifier. The removal of TSS is usually in the range of 75–85% and BOD_5 removal is 55–70% depending on the specific wastewater characteristics.

The use of metal salts for precipitation of phosphorus in suspended film biological systems is widely practised. The most common salts used are ferric chloride and sulfate and similar salts of aluminum. Pickle liquor, ferrous sulfate, and ferrous and aluminum chloride are also used. When metal salts are used, it may be necessary to provide additional alkalinity to the aeration basin.

Generally, the metal salts are used in excess molar ratio, i.e., moles Al:P or Fe:P. The excess metal salts form hydroxides of the metal and precipitate. The residuals produced are as follows:

a. $Al\,PO_4$ = 121/31 or 3.9 mg/mg P removed.
b. $Al\,(OH)_3$ = 77/26 or 3.0 mg/mg excess Al.
c. $Fe\,PO_4$ = 151/31 or 4.9 mg/mg P removed.
d. $Fe\,(OH)_3$ = 107/56 or 1.9 mg/mg excess Fe.

The residual soluble total phosphorus as a function of the molar ratio is approximately as follows:

Table 10
Typical Production of Primary and Primary-Chemical Biosolids[a]

Mode of operation	Dosage of chemical (mg/L)	Raw TSS removed (mg/L)	Raw BOD_5 removed (mg/L)	Chemical sludge produced (mg/L)
Plain sedimentation	–	120	60	–
Polymer added	0.5–3.0	150	90	–
CaO aided[b]	200	160	120	128
$FeCl_3$ (as Fe)[c]	12	160	120	47
Al_2SO_4 (as Al)[c]	12	160	120	46

Source: US EPA.

[a] Based on 200 mg/L·BOD_5, 200 mg/L TSS, and 10 mg/LTP in raw sewage; primary effluent ≤ 2 mg/L total phosphorus.

[b] Varies because of permanent hardness in the water, used 35 mg/L precipitated as $CaCO_3$.

[c] May require polymer addition to enhance clarification.

Molar ratio (metal: TP)	Residual soluble total phosphorus (mg/L)
1	2
1.5	1
2	0.3

The residual phosphorus at low levels is highly dependent on pH, and it might be more economical to increase the pH by adding alkalinity which will not produce a residual.

5. CHARACTERISTICS OF BIOSOLIDS

5.1. Specific Gravity and Volatility

The specific gravity of biosolids will be in part a function of the amount of grit and fine inert particles in the biosolids. These inorganic particles will have a specific gravity of 2.5–2.9. Where there is good degritting, the specific gravity of biosolids will have the volatile and specific gravities shown in Table 11. The specific gravities of dry solids are quite low and will vary depending on the source.

The specific gravity of fixed film biosolids is generally higher than that of WAS. This is evidenced by a lower sludge volume index (SVI) and generally higher settling rates. The specific gravity of the biosolids after anaerobic digestion will increase because of reduction of some of the hydrous fractions and the increased inert content. The solid matter in biosolids is made up of fixed and volatile solids. The specific gravity of all the solids can be computed as follows (6):

$$\frac{W_s}{S_s\rho_w} = \frac{W_f}{S_f\rho_w} + \frac{W_v}{S_v\rho_w} \tag{9}$$

where W_s is the weight of solids (lb); S_s is the specific gravity of solids; ρ_w is the density of water (lb/ft^3); W_f is the weight of fixed solids (lb); S_f is the specific gravity of fixed solids; W_v is the weight of volatile solids (lb); S_v is the specific gravity of volatile solids.

Equation (9) can be utilized to determine the overall specific gravity of total solids (fixed + volatile) in any of the following three convenient forms:

Table 11
Specific Gravity of Waste Biosolids

Sludge type	Volatility (%)	Range of specific gravity (g/mL)
RPS	75–80	1 + 0.010 (TSS %)–1 + 0.012 (TSS %)
WAS	80–85	1 + 0.007 (TSS %)–1 + 0.012 (TSS %)
TF and RBC	75–80	1 + 0.015 (TSS %)–1 + 0.025 (TSS %)
RPS + WAS	75–85	1 + 0.004 (TSS %)–1 + 0.006 (TSS %)

Source: US EPA.

RPS, raw primary sludge; WAS, waste-activated sludge; TF, trickling filter; RBC, rotating biological contactor.

$$W_s/S_s = W_f/S_f + W_v/S_v \qquad (10)$$

$$1/S_s = \text{Fixed solids fraction}/S_f + \text{volatile solids fraction}/S_v \qquad (11)$$

$$100/S_s = \text{Fixed solids (\%)}/S_f + \text{volatile solids (\%)}/S_v \qquad (12)$$

The specific gravity of the biosolids can be determined as follows:

$$(W_w + W_s)/S = W_w/S_w + W_s/S_s \qquad (13)$$

As $S_w = 1$, then Eq. (13) becomes:

$$(W_w + W_s)/S = W_w + W_s/S_s \qquad (14)$$

$$100/S = P_w + (100 – p_w)/S_s \qquad (15)$$

where S is the specific gravity of biosolids; W_w is the weight of water in biosolids (lb); P_w is the percentage of water in biosolids (%).

Rearranging Eqs. (14) and (15) yields the following expressions for computing S:

$$S = (W_w + W_s)/(W_w + W_s/S_s) \qquad (16)$$

$$S = 100/\{P_w + [(100 - P_w)/S_s]\} \qquad (17)$$

5.2. Preconcentration or Dewaterability of Biosolids

Raw primary biosolids are the easiest to thicken followed by fixed film biosolids. WAS is the most difficult to thicken, particularly if the SVI is high. Chemical biosolids produced from the addition of metal salts thicken similarly to WAS at a SVI = 100 mL/g, but they are more stable. Aging of biosolids after removal from the raw wastewater or the aerobic environment causes deterioration of the thickening quality.

The general experience in thickening biosolids is shown in Table 12. The results achievable in the primary clarifier are dependent on the clarifier design as explained earlier. Thickening increases the solids content of biosolids slurry by partial, but substantial, removal of the liquid phase. The purpose is to reduce the biosolids volume to be stabilized, dewatered, or hauled away (68). Figure 7 shows the importance of thickening before mechanical dewatering. Thickening can be accomplished by partial thickening in a primary or secondary clarifier, a gravity thickener, a dissolved air flotation thickener, a centrifuge, a gravity or low pressure belt press, or a rotary drum device.

Table 12
Thickening of Waste Biosolids

Type of sludge	TSS Concentration (%) Primary clarifier	 Flotation/DAF	 Gravity thickener	 Belt thickener[a]	 Centrifuge
RPS	5–7		8–10	9–12	9–12
WAS		3–5	2–2.5	4–6	4–6
FFS[b]		3–5	2.5–3	5–7	5–7
RPS + WAS	2.5–4	4–6	4–5	5–7	5–7
RPS + FFS	3–5	4–6	5–6	5–10	6–10

Source: US EPA.
[a] Polymers required.
[b] Fixed-film sludge.

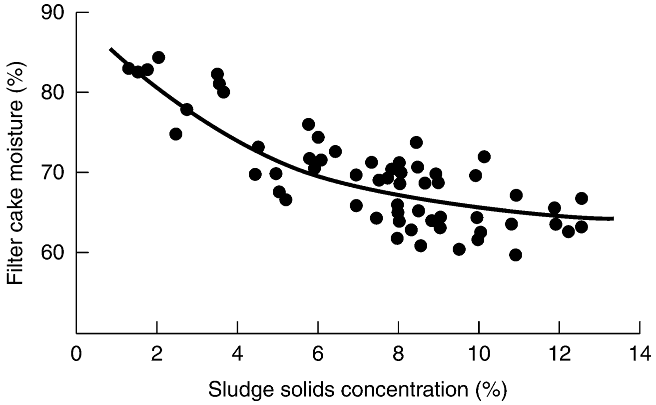

Fig. 7. Effects of feed solids on performance of a rotary vacuum filter (*Source:* US EPA).

The impact of decreasing the biosolids water content on the biosolids volume might be clarified by the following relationships. The weight of solids may be expressed by (6):

$$W_s = V \rho_w S P_s/100 \tag{18}$$

where W_s is the weight of dry solids (lb); V is the volume of biosolids (ft^3); ρ_w is the density of water (lb/ft^3); S is the specific gravity of biosolids; P_s is the percentage of solids present in the biosolids (%).

Applying Eq. (18) to the biosolids before and after dewatering yields the following expressions:

$$(Ws)_1 = V_1 \rho_w S_1 (P_s/100)_1$$

$$(W_s)_2 = V_2 \rho_w S_2 (P_s/100)_2$$

As there is no change in the amount of dry solids,

$$(W_s)_1 = (W_s)_2:$$

$$V_1 \rho_w S_1 (P_s/100)_1 = V_2 \rho_w S_2 (P_s/100)_2$$

$$V_1/V_2 = (S_2/S_1)\,(P_s)_2/(P_s)_1$$

However, the term (S_2/S_1) is near unity and for all practical purposes might be neglected. Hence:

$$V_1/V_2 = (P_s)_2/(P_s)_1 \qquad (19)$$

So, it is clear that when water is removed from the biosolids the volume of biosolids varies inversely with its solids content. For example, reducing the biosolids moisture content by just 1% point from 99 to 98% (i.e., increasing the solids content from 1 to 2%) reduces the biosolids volume to half its original value.

Gravity thickening of raw or digested primary biosolids is almost always an efficient and economical process. Anaerobically digested primary biosolids are normally thickened by gravity in the secondary digester. The use of primary basins to capture and to thicken both wastewater influent and recirculated WAS solids, might not always be a cost effective and efficient practice in larger plants. The WAS solids might not resettle well in hydraulically overloaded or septic primary tanks. Hence, this practice results in the production of more WAS because of to an increased solids load on the aeration system. Poorer thickening results when the primary basins are used to concentrate the WAS solids, particularly if the bottom configuration is not conducive to thickening.

The use of gravity thickeners for both RPS + WAS has had mixed results. Most of the poor results can be traced to one or more of the following causes:

a. RPS + WAS feed concentration is >0.5% TSS.
b. RPS is very septic.
c. WAS is > RPS fraction.
d. Secondary dilution water is inadequate.
e. Floor slope is <2.5:12, causing excessive solids retention.
f. Biosolids not removed continuously.

Properly designed and operated gravity thickeners work effectively on mixtures of RPS and WAS throughout the United States. Misusing them as biosolids storage zones causes operator grief. If storage is necessary, it must be placed after the gravity thickeners. The use of other thickening methods such as dissolved air flotation, basket or solid bowl centrifugation, low pressure belt filtration, and the rotary drum system has increased because these methods can also give reliable and effective results when thickening WAS.

5.3. Particle Surface Charge and Hydration

Biosolids particles have a negative surface charge and try to repel each other as they are brought together. Additionally, biosolids particles weakly attract water molecules to their surface (hydration) either by weak chemical bonding or by capillary action. Although the water is only weakly held at the particle surface, it does resist thickening and interferes with dewatering. Chemical conditioning is used to overcome the effects of surface charge and surface hydration. Typical chemicals are organic polymers; lime, ferric chloride and other metallic salts. Generally, they act by reducing or eliminating the repulsive force, thus permitting the particles to come together or flocculate. Water can be more readily removed at a higher rate during the subsequent mechanical dewatering.

Table 13
Biosolids Dewatering as a Function of Particle Size

Mean diameter (µm)	Specific resistance (s^2/g)	Particles (% of total)
Original, unfractionated sample	10.4×10^9	–
>100	2.3×10^9	10.2
5–100	4.6×10^9	75.5
1–5	13.8×10^9	8.5
<1	–	5.9

Source: US EPA.

5.4. *Particle Size*

Generally, particle size is recognized as a very important factor influencing dewaterability. As the average particle size decreases, the surface area and surface-to-volume ratio for a given biosolids mass increase. The effects of increasing the surface area include:

a. Higher repulsion between the particles resulting from the larger area of negatively charged surface.
b. Higher attraction of water to the particle surface because of more sites for chemical joining.

Particle size is influenced by both the biosolids source and prior treatment. Primary biosolids, in addition to containing more inorganic and fibrous materials, have a larger average particle size than secondary biosolids. This is because fine suspended and colloidal solids tend to pass through the primary clarifier. Biosolids particles passing the primary clarifier are then removed in the secondary clarifier along with the less dense, flocculated cellular material that is created during biological treatment. The activated sludge process, in addition to removing most of dissolved BOD, functions to capture, remove, and hence recover most of these residual materials by biocoagulation and flocculation. As a result, activated sludge (biosolids) is finer than primary biosolids. Normally, it consist of 60–90% or more cellular organic material and contains a very large amount of water.

Individual particles of activated sludge are usually aggregated to an extent through bioflocculation. Table 13 shows the relative difficulty of removing water from an unflocculated primary digested biosolids containing various particle size fractions. As can be seen, the specific resistance to filtration of the unfractionated biosolids is dominated by the specific resistance of material under 5 µm in size, eventhough this material constitutes only about 14% by weight of the total solids (69). Specific resistance is, in effect, a measure of the relative dewaterability of biosolids. The lower the specific resistance, the higher is the biosolids' dewaterability. Specific Resistance has been defined as the pressure required to produce a unit rate of flowthough a cake having a unit weight of dry solids per unit area when the viscosity of the liquids is unity. Specific values are determined from laboratory filtration experiments:

$$r = 2\mathrm{b}\ (\Delta P)\ A^2/\mu\ \omega \qquad (20)$$

where r is the specific resistance (s^2/lb); b is the slope of a plot of t/V vs V; t is the time (s); V is the filtrate volume (ft^3); ΔP is the pressure differential across the biosolids

Table 14
Specific Resistance of Various Types of Biosolids

Type of sludge	Specific resistance (s^2/g)
Raw	$10\text{–}30 \times 10^9$
Raw (coagulated)	$3\text{–}10 \times 10^7$
Digested	$3\text{–}30 \times 10^9$
Digested (coagulated)	$2\text{–}20 \times 10^7$
Activated	$4\text{–}12 \times 10^9$

Source: US EPA.

cake (lb/in^2); A is the filter area (ft^2); μ is the absolute viscosity (lb/s-ft); ω is the weight of solids deposited per unit volume of filtrate (lb/ft^3).

Table 14 contains typical specific resistance values for different types of biosolids, both chemically treated and untreated. As the maximum specific resistance for feasible mechanical dewatering is normally quoted at $\leq 10 \times 10^7$ s^2/g, none of these biosolids would be readily dewaterable. Table 14 shows that specific resistance values can vary significantly. Experience indicates that properly conditioned RPBs are almost always the most readily dewatered, followed by well-conditioned digested primary biosolids and then activated sludges, in increasing order of difficulty.

Biosolids stabilization by aerobic and anaerobic processes results in the destruction of a portion of the organic matter and the production of hydrous particles, which are more difficult to dewater. However, a significant portion of the original hydrous biosolids is also destroyed in the stabilization process. The consequence is that the residual digested biosolids are sometimes more difficult to dewater, sometimes easier. But, in any case, the quantity is reduced 30–40% from the raw state.

5.5. Compressibility

If biosolids particles were idealized incompressible solids, the solids would not deform, and the void space between the particles would remain constant during mechanical dewatering. In such an ideal situation, resistance to filtration would be proportional to biosolids depth, and there would be no increase in resistance to filtration as dewatering progresses. Unfortunately, biosolids particles are practically always hydrophilic and compressible to a degree, which results in particle deformation and a reduction in the void area between particles. This reduction in void volume inhibits the movement of water through the compressed portion of the biosolids cake, and reduces the rate of dewaterability. The compressibility coefficient (S), is an empirical measure of the effect of pressure on the permeability of biosolids cake:

$$r = K_c\,(\Delta P)^{S} \tag{21}$$

where r is the specific resistance (s^2/lb); K_c is the cake constant; ΔP is the pressure differential across the biosolids cake, ($lb/in.^2$); S is the compressibility coefficient. The value of S can be determined by the analysis of the specific resistance laboratory filtration data obtained at various pressure differentials. The lower is the value of S the less is the compressibility of the biosolids.

Proper conditioning improves dewaterability primarily by producing a flocculent matrix of solids in relatively clear water before filtration. When this matrix is deposited on a filtering medium, the bulk cake retains a substantial porosity. However, too high a pressure drop across the biosolids floc will trigger the conditioned biosolids cake to collapse, and will result in a decreased filtration rate. The net result of conditioning is quicker removal of water, principally, because of the higher rate of water removal at the start of the filtration cycle.

5.6. Biosolids Temperature

As biosolids temperature increases, the viscosity of the water present in the biosolids mass decreases. Viscosity is particularly important in centrifuge dewatering because sedimentation is the key component of the process (70).

5.7. Ratio of Volatile Solids to Fixed Solids

Biosolids tend to dewater better as the percentage of fixed solids increases. In fact some high-g centrifuge manufacturers use the percentage of fixed solids as a key parameter in sizing equipment. The biosolids cake from centrifugal dewatering of an anaerobically digested mixture of primary biosolids and WAS, shows a positive change of 5% in its solids concentration, when the percentage of volatile solids in it decreases from 70 to 50%. However, because digestion also produces smaller particles, the higher surface area results in more moisture. The earlier-mentioned approximation of volatile content to cake solids must be carefully used and should be pretested whenever possible.

5.8. Biosolids pH

Biosolids pH affects the surface charge on biosolids particles. Hence, pH will influence the type of polymer to be used for conditioning. Generally anionic polymers are the most useful when the biosolids are lime conditioned and have a high pH, whereas cationic polymers are most suitable at a pH slightly higher or below neutral. In some cases, cationic polymers can be effective up to pH 12.0 and have been used for lime stabilized biosolids.

5.9. Septicity

Septic biosolids are more difficult to dewater and require higher dosages of chemical conditioners than fresh biosolids. This phenomenon has been experienced at many locations and is the most likely because of a reduction in the size of biosolids particles, to the generation of gases that remain entrained in the biosolids, and to the change in surface characteristics created by bioconversion. Wetter cake and lower biosolids production are common results from dewatering septic biosolids. For this reason raw biosolids storage should be minimized as an operating practice.

5.10. Trace Elements and Heavy Metals

It is a basic principle in chemistry that elements are not created or destroyed but chemically recombined. Therefore the mass of each element entering a treatment plant fixes the mass that either accumulates within the plant or leaves it. The mass

Table 15
Metal Content in Wastewater Biosolids

Metal	Dry biosolids (range mg/kg)	Median (mg/kg)
Arsenic	1.1–230	10
Cadmium	1–3410	10
Chromium	10–99,000	500
Cobalt	11.3–2490	30
Copper	84–17,000	800
Iron	1000–154,000	17,000
Lead	13–26,000	500
Manganese	32–9870	260
Mercury	0.6–56	6
Molybdenum	0.1–214	4
Nickel	2–5300	80
Selenium	1.7–17.25	5
Tin	2.6–329	14
Zinc	101–49,000	1700

Source: US EPA.

leaving the plant does so in gaseous emissions, effluent, a special concentrated stream, or biosolids.

Trace elements are present in industrial process waste, industrial waste spills, domestic water supply, feces and urine, and detergents. Additional trace elements are derived from (4):

a. Chemicals in photographic solutions, paints, hobby plating supplies, dyes, and pesticides used in households and commercial enterprises.
b. Stormwater inflow.
c. Corrosion of water piping, which contributes zinc, cadmium, copper, and lead.
d. Chemicals used in wastewater treatment, biosolids conditioning, and so on.

Typical heavy metals, trace elements found in biosolids, vary widely as shown in Table 15. High concentrations of such toxic elements may limit the extent of biosolids utilization by application to land (71).

6. EXAMPLES

6.1. Example 1: Determination of Biosolids Volume

Assume activated sludge plant with primary clarifiers:

a. **Sludge type.** Primary biosolids (P) + WAS.
b. **Biosolids quantities.** P—150 kg solids/million liters (ML) treated [1250 lb/million gallons (MG)].

WAS—90 kg solids/ML treated (750 lb/MG).

c. **Solids concentrations.** P—5% from primary clarifier.

WAS—0.5% from secondary clarifier.

Thickened WAS—4% from dissolved air flotation thickener.

d. **Assume solids specific gravity.** 1.4 For primary biosolids and 1.25 for WAS.

Determine biosolids volumes before anaerobic digestion.

Solution

Biosolids specific gravity:

$$S = 100/\{P_w + [(100 - P_w)/S_s]\} \tag{17}$$

Primary biosolids specific gravity:

$$S = 100/\{95 + [(100 - 95)/1.4]\} = 1.01$$

Thickened WAS specific gravity:

$$S = 100/\{96 + [100 - 96)/1.25]\} = 1.01$$

Therefore, combined P + WAS has $S = 1.01$

Biosolids volume:

$$W_s = V \rho_w S P_s/100 \tag{18}$$

$$V = W_s (100)/\rho_w S P_s$$

$$V = [150 (100)/(1) (1.01) (5)] + [90 (100)/(1) (1.01) (4)]$$

$$= 2970 + 2230 \text{ L/ML treated.}$$

$$= 5200 \text{ L/ML treated (5200 gal/MG).}$$

6.2. Example 2: Determination of Solids Content After Digestion

Assume that anaerobic digestion of the biosolids in example 1 destroys 50% of the volatile solids. Also assume 70% of the solids are volatile. Determine the solids content after digestion.

Solution

a. Solids content before digestion = (100) (150 + 90)/(5200) (1) (1.01) percent = 4.6% solids.
b. Mass of solids before digestion: 150 + 90 = 240 kg/ML.
c. Solids destroyed = (240 kg/ML) (0.7) (0.50) = 84 kg/ML (700 lb/MG).
d. Solids remaining = 240 kg/ML − 84 kg/ML = 156 kg/ML (1300 lb/MG).
e. Solids content after digestion = (100) (156 kg/ML)/(5200) (1) (1.01) percent = 3% solids.

6.3. Example 3: Determination of Biosolids Production

This example illustrates the use of yield factors and decay factors. Figure 8 shows a flow diagram for a hypothetical plant. The problem is to prepare an initial estimate of the loading to the WAS thickener. Table 16 contains information required for this calculation, including average and maximum day loadings and activated sludge operating characteristics.

It is assumed that the thickener in this example will have to handle the maximum-day WAS production. Peak loadings of shorter duration than the maximum day production will be handled by storing the added suspended solids in the aeration basins. For the purposes of this example, the biosolids treatment processes such as digestion, dewatering, disinfection, thermal conditioning, and chemical conditioning have not been identified. Depending upon the selection and design of the biosolids treatment processes, the recycle loads from such processes could have a significant effect upon the quantities of WAS and primary biosolids that must be processed. When they are known, the degradable

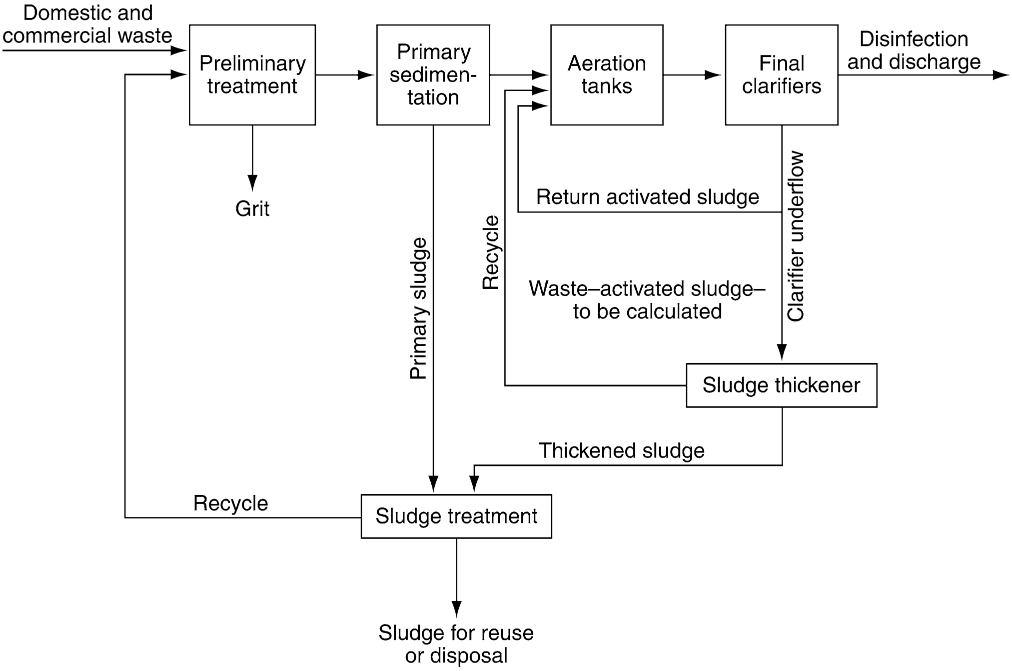

Fig. 8. Schematic for biosolids quantity example (*Source:* US EPA).

organics (BOD_5), and nonvolatile fractions of the recycle streams should be added to the substrate removal (S_r), and nonvolatile suspended solids (I_{NV}) factors. Subsequent calculations in Eqs. (1) and (2) are for the purposes of obtaining biosolids mass balance, which includes the effect of recycle streams.

Solution

1. Determine BOD_5 load to the activated sludge process.
   a. Average day BOD_5 load.

$$5 \text{ MGD} \times (8.34 \text{ lb/MG/1 mg/L}) \times 190 \text{ mg/L} \times (1 - 0.35) = 5150 \text{ lb/d.}$$

   b. Maximum day BOD_5 load.

$$9.5 \text{ MGD} \times (8.34 \text{ lb/MG/1 mg/L}) \times 160 \text{ mg/L} \times (1 - 0.25) = 9510 \text{ lb/d.}$$

2. Determine M, the mass of microorganisms.
   a. Average day.

$$F/M = (BOD_5 \text{ applied/d})/(\text{VSS in system}) = 0.3.$$

$$M = 5150/0.3 = 17{,}170 \text{ lb VSS.}$$

   b. Maximum day.

$$F/M = 0.5.$$

$$M = 9150/0.5 = 19{,}020 \text{ lb VSS.}$$

3. Determine Y, the gross yield coefficient and k_d, the decay coefficient.

Table 16
Design Data for Biosolids Production Example

Description	Value
Influent flow [MGD(m^3/d)]	
Average day	5 (18,900)
Maximum day	9.5 (36,000)
Influent BOD_5 (mg/L)	
Average day	190
Maximum day	160
Influent suspended solids (mg/L)	
Average day	240
Maximum day	190
BOD_5 removal in primary sedimentation (%)	
Average day	35
Maximum day	25
Suspended solids removal in primary sedimentation	
Average day	65
Maximum day	50
Sludge thickener capture efficiency	
Average day (%)	95
Maximum day (%)	85
Food-to-microorganism ratio[a]	
Average day	0.3
Maximum day	0.5
Temperature of wastewater	
Average day, °F (°C)	65 (18)
Maximum day °F (°C)	50 (10)
Dissolved oxygen in aeration tanks	
Average day (mg/L)	2.5
Minimum day (mg/L)	2
Control (automatic)	–
Effluent limitations (30-d average)	
BOD_5 (mg/L)	30
Suspended solids (mg/L)	30
Usable test data for solids production	None[b]

Source: US EPA.

[a] lb (kg) BODs applied daily/lb (kg) mixed liquor VSS.

[b] Data from other plants must be used.

1 MGD = 3785 m^3/d.

Note: Maximum day influent BOD_5 and suspended solids concentrations reflect a dilution from average day data because of the higher flow.

No test data are available for this waste, so estimates must be made from tests on other wastes: For average conditions, use the data for full-scale PE from Table 6:

$$Y = 0.67 \text{ lb (kg) VSS formed/lb (kg) } BOD_5 \text{ removed; } k_d = 0.06 \text{ d}^{-1}.$$

For maximum conditions, use minimum temperature of 36°F (10°C), which produces the maximum Y value. Apply the temperature correction, which increases Y by 26%:

$Y = 0.67 \times 1.26 = 0.84$; do not adjust k_d.

4. Determine S_r (substrate removal) in units to match Y.
   a. Average daily substrate removal.
      i. BOD_5 applied = 5150 lb/d.
      ii. Effluent BOD_5 (assume 10 mg/L) = 420 lb/d.
      iii. BOD_5 removed = 5150 – 420 = 4730 lb/d.
   b. Maximum daily substrate removal.
      i. BOD_5 applied = 9510 lb/d.
      ii. Effluent BOD_5 (assume 10 mg/L) = 790 lb/d.
      iii. BOD_5 removed = 9510 – 790 = 8720 lb/d.

Note: Allow 10 mg/L for effluent BOD_5, even though the plant is permitted to discharge 30 mg/L. Activated sludge plants can often attain 10 mg/L effluent BOD_5. Biosolids capacity should be provided for the biosolids produced under such conditions.

5. Determine P_X the biological solids production.

$$P_X = (Y)\,(S_r) - (k_d)\,(M) \qquad (1)$$

   a. Average day.

$$P_X = (0.67)\,(4730) - (0.06)\,(17{,}170) = 2140 \text{ lb VSS produced/day.}$$

   b. Maximum day.

$$P_X = (0.84)\,(8720) - (0.06)\,(19{,}020) = 6184 \text{ lb VSS produced/day.}$$

6. Compute I_{NV} (nonvolatile suspended solids fed to the activated sludge process).
   a. Average daily input of nonvolatile suspended solids.

$$5 \text{ MGD} \times 8.34 \times 240 \text{ mg/L } (1 - 0.65)\,(0.25) = 880 \text{ lb/d.}$$

   b. Maximum daily input of nonvolatile suspended solids.

$$9.5 \text{ MGD} \times 8.34 \times 190 \text{ mg/L } (1 - 0.50)\,(0.25) = 1800 \text{ lb/d.}$$

7. Compute E_T (effluent suspended solids).
   a. Average day.

$$5 \text{ MGD} \times 8.34 \times 10 \text{ mg/L} = 420 \text{ lb/d.}$$

   b. Maximum day.

$$9.5 \text{ MGD} \times 8.34 \times 10 \text{ mg/L} = 790 \text{ lb/d.}$$

8. Compute WAS_T production.

$$WAS_T = P_X + I_{NV} - E_T \qquad (2)$$

   a. Average day.

$$WAS_T = 2140 + 880 - 420 = 2600 \text{ lb TSS/d (1180 kg/d).}$$

   b. Maximum day.

$$WAS_T = 6184 + 1880 - 790 = 7274 \text{ lb TSS/d (3302 kg/d).}$$

9. Compute inventory reduction allowance.

$$\text{Inventory reduction allowance} = (0.02)\,(17{,}170) = 343 \text{ lb/d (156 kg/d).}$$

In the present case, the inventory reduction allowance can be small. Allow 2% of *M*/d. The 343 lb/d computed here is smaller than the difference between the average and maximum WAS production (Step 8); therefore, if capacity is provided for maximum solids production, then there will be ample capacity for inventory reduction. It is not necessary to reduce inventory during peak loads.

6.4. Example 4: Interaction of Yield Calculations and the Quantitative Flow Diagram

The example just presented demonstrates a technique for calculating solids production on a once-through basis; that is, any solids associated with recycle streams were not considered in the calculation. The quantitative flow diagram (QFD) considers the effects of recycle streams. Before the QFD can be constructed for biological treatment processes, an estimate of net solids destruction or synthesis must first be made. The relationship between solids entering and leaving the biological unit is established through the parameter X_D, which is defined as net solids destruction per unit of solids entering the biological unit. The data and calculations from the previous design example allow an initial estimate of X_D to be made.

Solution

For the average flow:

a. Solids leaving the biological unit = $P_X + I_{\mathrm{NV}} = 2140 + 880 = 3020$ lb/d.
b. Solids entering the biological unit are equal to solids in the PE, which can be calculated from the data on Table 6. PE solids = $(1 - 0.65)\ (240)\ (8.34)\ (5) = 3503$ lb/d.
c. Net solids destruction = solids in – solids out = $3503 - 3020 = 483$ lb/d (219 kg/d).
d. $X_D = 483/3503 = 0.138$.

For maximum day flows:

a. Solids leaving the biological unit = $6184 + 1880 = 8064$ lb/d (3661 kg/d).
b. Solids entering the biological unit = $(1 - 0.50)\ (190)\ (8.34)\ (9.5) = 7527$ lb/d (3147 kg/d).
c. Net solids destruction = $8064 - 7527 = 537$ lb/d (244 kg/d).
d. $X_{D\mathrm{max}} = 537/7527 = 0.07$.

Once X_D is known, the QFD calculation can be undertaken. After the QFD calculation is completed, the designer may wish to make new estimates of P_X and I_{NV} based on information derived from the QFD calculation. For example, if the QFD calculation shows that recycle loads are substantial, then the designer might wish to modify estimates of S_r and I_{NV} and calculate new values of P_X and I_{NV}.

NOMENCLATURE

A	Filter area (ft^2)
A_m	Total media surface area in reactor (ft^2 or m^2)
b	Slope of a plot of t/V vs V
C_2	Coefficient to match units of S_r and "F" in F/M; if S_r is BOD_5 removed (influent – effluent), then C_2 is BOD_5 removal efficiency, about 0.9
E_T	Effluent suspended solids (lb/d or kg/d)
F/M	Food-to-microorganism ratio = BOD_5 applied daily/VSS (mass) in system
I_{NV}	Nonvolatile suspended solids fed to the process (lb/d or kg/d)
k_d	Decay coefficient (activated sludge) (d^{-1})
$k'd$	Decay coefficient (trickling filter) (d^{-1})

K_c	Cake constant
M	System inventory of microbial solids (VSS) (lb or kg)
P_s	Percentage of solids present in the biosolids (%)
P_w	Percentage of water in biosolids (%)
P_X	Net growth of biological solids (VSS) (lb/d or kg/d)
P_x/M	Net growth rate = $1/\theta_m$ d^{-1}
ΔP	Pressure differential across the biosolids cake (lb/in^2)
r	Specific resistance (s^2/lb)
S	Compressibility coefficient
S	Specific gravity of biosolids
S_f	Specific gravity of fixed solids
S_r	Substrate (e.g., BOD$_5$) removed (lb/d or kg/d)
S_s	Specific gravity of solids
S_v	Specific gravity of volatile solids
S_r/M	lb (kg) BOD$_5$ removed per day/lb (kg) VSS
t	Time (s)
V	Filtrate volume (ft^3 or m^3)
V	Volume of biosolids (ft^3 or m^3)
W_f	Weight of fixed solids (lb or kg)
W_s	Weight of solids (lb or kg)
W_w	Weight of water in biosolids (lb or kg)
W_v	Weight of volatile solids (lb or kg)
WAS_T	Waste-activated sludge production (lb/d or kg/d)
WTFB	Waste trickling filter biosolids production (lb/d or kg/d)
Y	Gross yield coefficient (activated sludge) (lb/lb or kg/kg)
Y'	Gross yield coefficient (trickling filter) (lb/lb or kg/kg)
Y_{obs}	Net yield coefficient = lb(kg) VSS produced /lb(kg) substrate (e.g., BOD$_5$) removed
θ_m	M/P_X = Sludge age (d)
ρ_w	Density of water (lb/ft^3 or kg/m^3)
μ	Absolute viscosity (lb/s-ft or kg/s-m)
ω	Weight of solids deposited per unit volume of filtrate (lb/ft^3 or kg/m^3)

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2

Gravity Thickening

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

1.1. General

Thickening is defined as removal of water from sludge to achieve a reduction in moisture content of slurries. The resulting material is still fluid. Thickening is used at most wastewater treatment plants, as an economic measure, to reduce the volume of sludge or for greater efficiency in subsequent processes. Sludges are thickened primarily to decrease the capital and operating costs of subsequent sludge processing steps by substantially reducing the volume. Thickening from 1 to 2% solids concentration, for example, halves the sludge volume. Further concentration to 5% solids, reduces the volume to one-fifth of its original volume.

Depending on the process selected, thickening might also provide the following benefits:

a. Sludge blending.
b. Sludge flow equalization.
c. Sludge storage.
d. Grit removal.
e. Gas stripping.
f. Clarification.

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The purpose of this chapter is to provide the reader with rational design and operating information on which to base decisions about a cost-effective thickening process. Thickening is only one part of wastewater biosolids treatment and disposal system and must be integrated into the overall treatment process, so that performance for both liquid and solid treatment is optimized and total cost is minimized (1–5).

Gravity or flotation thickeners are normally used to accomplish thickening; centrifuges have also been used as sludge thickeners. Gravity thickening is the subject at hand while air flotation and centrifugation are presented in other chapters.

1.2. Gravity-Thickening

Gravity-thickening is the most common process currently used for dewatering and for the concentration of sludge before digestion. Gravity-thickening is essentially a sedimentation process similar to the process that occurs in all settling tanks. The process is simple and is the least expensive of the available thickening processes (6,7).

Gravity-thickening might be classified as plain settling and mechanical thickening. Plain settling usually results in the formation of scum at the surface and stratification of sludges near the bottom. Sludges from secondary clarifiers usually cannot be concentrated by plain settling. Gentle agitation is usually employed to stir the sludge, thereby opening channels for water to escape and promoting densification. A common mechanical thickener consists of a circular tank equipped with a slowly revolving sludge collector. Primary and secondary sludges are usually mixed before thickening. A ratio of secondary sludge to primary sludge of 8:1 or more is recommended to ensure aerobic conditions in the thickener. Chlorine has been used to prevent sludge septicity and gasification, which interfere with optimum solids concentration of organic materials. A chlorine residual of 0.5–1 mg/L in the thickening tank prevents such problems. Organic polyelectrolytes (anionic, nonionic, and cationic) have been used successfully to increase the sludge settling rates, the overflow clarity, and the allowable tank loadings.

1.3. Process Evaluation

Although it is good design practice to pilot thickening equipment before designing a facility, pilot testing does not guarantee a successful full-scale system. Designers must be aware of the difficulties involved in scale-up and the changing character of wastewater sludge and allow for the changing parameters in the design.

The main design variables of any thickening process are:

a. Solids concentration and volumetric flow rate of the feed stream.
b. Chemical demand and cost if chemicals are used.
c. Suspended and dissolved solids concentrations and volumetric flow rate of the clarified stream.
d. Solid concentration and volumetric flow rate of the thickened sludge.

Specific design criteria for selection of a thickening process can also be dependent on the chosen downstream process train. Another important consideration is the operation and maintenance (O/M) cost and the variables affecting it. In the past, O/M costs have not been given enough attention. However, this changed as US Environmental Protection Agency (US EPA) began to implement its Operations Check List in all phases of the Construction Grants Program (5).

Finally, thickening reliability is important for successful plant operation. A reliable thickening system is needed to maintain the desired concentration and relatively uninterrupted removal of sludge from a continuously operated treatment plant. Sludges are being generated constantly, and if they are allowed to accumulate for a long time, the performance of the entire plant will be degraded.

1.4. Types and Occurrence of Thickening Processes

As mentioned earlier Thickening is accomplished in sedimentation basins and in separate gravity, flotation, and centrifugal thickeners; and in miscellaneous facilities such as secondary anaerobic digesters, elutriation basins, and sludge lagoons.

2. SEDIMENTATION BASINS

2.1. Primary Sedimentation

A primary clarifier can be used as a thickener under certain conditions. Primary sludge thickens well, provided the sludge is reasonably fresh, solids of biological origin (e.g., waste activated sludge [WAS]) are kept to a minimum, and the wastewater is reasonably cool. If sludge of 5–6% solid content is to be recovered from a primary sedimentation system, it is essential that the sludge transport facilities be designed to move those solids. This will require short suction piping, adequate net positive suction heads on the primary sludge pump, suction-sight glass inspection piping, and a positive means of ascertaining the quantity pumped and the concentration of the slurry (8).

2.2. Secondary Sedimentation

Thickening in secondary or intermediate clarifiers has not been successful in the past because biological sludges are difficult to thicken by gravity. Thickening has been improved by using side water depths of 14–16 ft (4–5 m) suction sludge withdrawal mechanisms rather than plow mechanisms, and gentle floor slopes, for example, 1:12. Although thickening within a sedimentation basin can be beneficial under certain conditions, separate thickening is usually recommended.

3. GRAVITY THICKENERS

3.1. Introduction

Separate, continuously operating gravity-thickening for municipal wastewater sludges was conceptualized in the early 1950s (9). Until that time, thickening had been carried out within the primary clarifier. Operating problems such as floating sludge, odors, dilute sludge, and poor primary effluent led to the development of the separate thickening tank. Now, gravity-thickeners became the most commonly used sludge-concentrating device; however, their use is being challenged by other thickening processes. The advantages and disadvantages of gravity thickeners compared with other thickeners are as follows (4).

The advantages of gravity-thickening are:

a. Provides the greatest sludge storage capabilities.
b. Requires the least operational skill.
c. Provides the lowest operation (especially power) and maintenance cost.

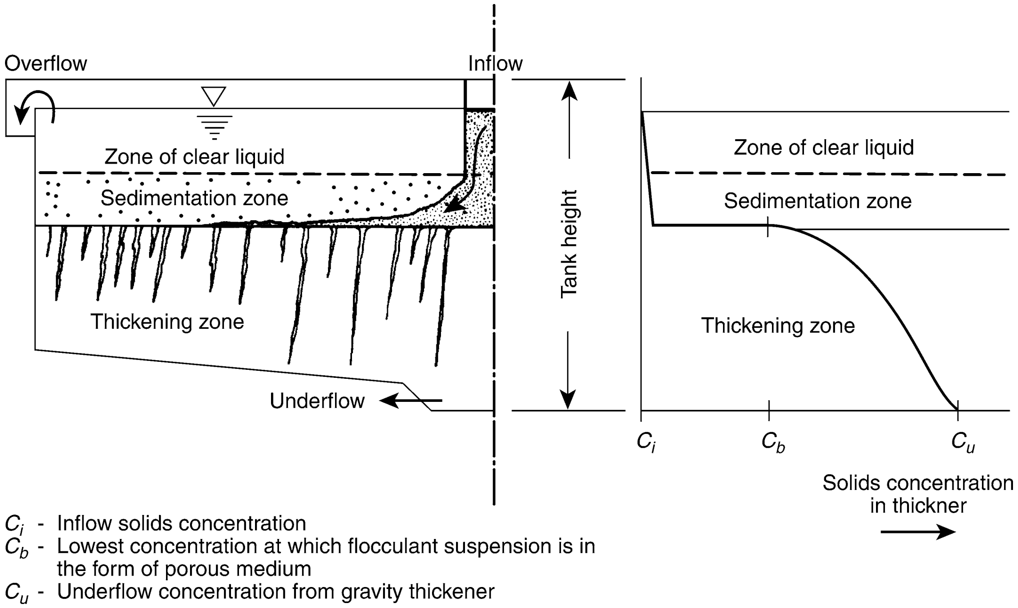

Fig. 1. Concentration profile of municipal sludge in a continuously operating thickener.

The disadvantages include the following aspects:

a. Requires the largest land area.
b. Contributes to the production of odors.
c. For some sludges, solid/liquid separation can be erratic.
d. For WAS, it produces the thinnest least concentrated sludge.

3.2. *Theory*

Since the early work of Coe and Clevenger (10), understanding of gravity-thickening has slowly improved (11–18). The key to understanding the continuous gravity-thickening process is recognition of the behavior of materials during thickening.

Coarse minerals thicken as particulate (nonflocculent) suspensions. However, municipal wastewater sludges are usually flocculent suspensions that behave differently (19). Only a short descriptive summary of thickening theory will be given here. Detailed, comprehensive analysis of gravity-thickening theory for municipal wastewater sludges is beyond the scope of this chapter; those desiring such detail should consult the works of Keinath et al. (20), Bustos et al. (17), and Shammas et al. (18).

Figure 1 shows a typical solids concentration profile for municipal wastewater sludges within a continuously operating gravity thickener. Sludge moving into the thickener partially disperses in water in the sedimentation zone and partially flows as a density current to the bottom of the sedimentation zone. The solid phase of the sludge, both dispersed and in the density current, creates flocs that settle on top of the thickening zone. Flocs in the thickening zone lose their individual character. They have mutual contacts and thus become a part of the matrix of solids compressed by the pressure of the overlying solids. The displaced water flows upward through channels in the solids matrix.

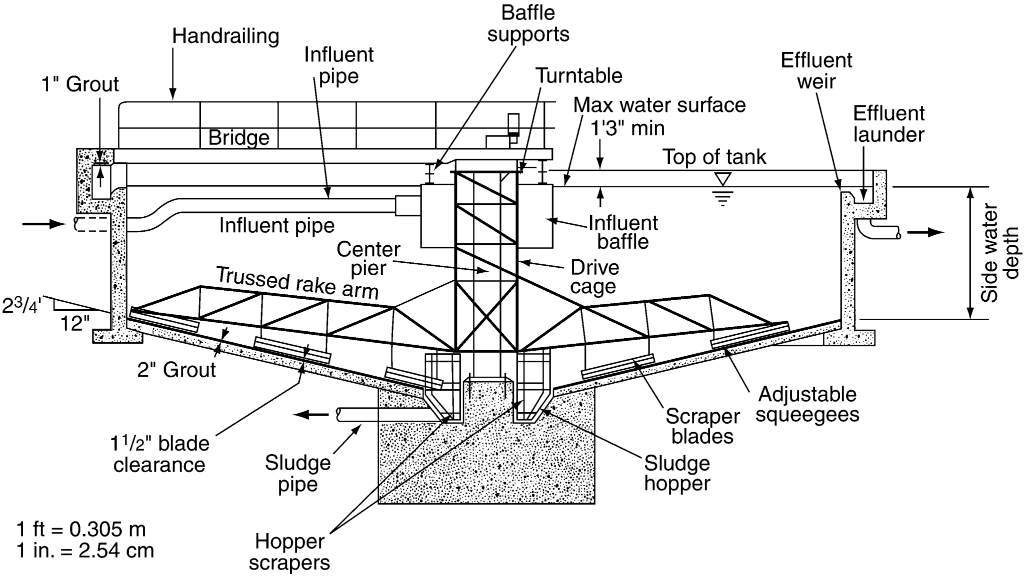

Fig. 2. Cross-sectional view of a circular gravity thickener.

Generally, in decision making about thickener size, the settling process in the sedimentation zone as well as the consolidation process in the thickening zone should be evaluated; whichever process (sedimentation or thickening) requires greater surface area dictates the size of the thickener. For municipal wastewater sludges, the thickening zone area required is almost always more than that for the sedimentation zone.

3.3. System Design Considerations

Circular concrete tanks are the most common configuration for continuously operating gravity thickeners, although circular steel tanks and rectangular concrete tanks have also been used. Figure 2 shows a cross-sectional view of a typical circular gravity thickener.

The following parameters should be evaluated for every gravity thickener:

a. Minimum surface area requirement.
b. Hydraulic loading.
c. Drive torque requirement.
d. Total tank depth.

Floor slope and several other considerations such as automation (21), strategy (22), and future perspectives (23) will also influence the final design of the gravity thickener.

3.3.1. Minimum Surface Area Requirement

If sludge from the particular facility is available for testing, the required surface area can be found by using a settling column, developing a settling flux curve, and calculating the critical flux (mass loading [lb/ft^2/h]) for that particular sludge (5,18,20). However, in most cases the sludge to be thickened is not available, and the designer must resort to other methods.

Table 1 provides criteria for calculating the required surface area when test data is not available and pilot plant work is not reasonable. The designer must specify the sludge type (for mixtures, the approximate proportions should be known), the range of solids concentration that is expected in the thickener inflow, and the underflow concentration required for downstream processing. A design example given at the end of the chapter will illustrate the use of Table 1 in sizing gravity thickeners.

3.3.2. Hydraulic Loading

Hydraulic loading is important for two reasons. First, it is related to mass loading. The quantity of solids entering the thickener is equal to the product of the flow rate and solids concentration. As there are definite upper limits for mass loading, therefore there will be some upper limit for hydraulic loading. Second, high hydraulic loading causes excessive carryover of solids in the thickener effluent.

Typical maximum hydraulic loading rates of 25–33 gal/ft^2/h (1200–1600 L/m^2/h) have been used in the past but mainly for primary sludges. For sludges such as WAS or similar types, much lower hydraulic loading rates, 4–8 gal/ft^2/h (200–400 L/m^2/h) are more applicable (24–27). Table 2 gives some typical operating results (24,28–32). Note that the hydraulic loading rate in gal/ft^2/h can be converted to an average upward tank velocity in feet per hour by dividing with 7.48.

Using the typical maximum hydraulic loading rates mentioned earlier, maximum velocities for primary sludges are 3.3–4.4 ft/h (1–1.3 m/h) and for WAS are 0.5–1.1 ft/h (0.2–0.3 m/h). Several researchers have related overflow rates to odor control, but odor is resulting from excessive retention of solids and can be better controlled by removing the thickened sludge from the thickener at an increased frequency.

3.3.3. Drive Torque Requirement

Sludge on the floor of a circular thickener resists the movement of the solids rake and thus produces torque. Calculation of torque for a circular drive unit is based on the simple cantilevered beam equation represented by Eq. (1):

$$T = WR^2 \tag{1}$$

where T is the torque (ft-lb), W is the uniform load (this is sludge specific [lb/ft]) (*see* Table 3), R is the tank radius (ft).

Note that there are several levels of torque, which must be specified for a circular gravity thickener. The following list defines the various torque conditions applicable to circular gravity thickeners (33).

a. Running torque—this is the torque value calculated from Eq. (1).
b. Alarm torque—torque setting, normally 120% of running, which tells the operator that there is something wrong.
c. Shut-off torque—torque setting, normally 140% of running, which would shut down the mechanism.
d. Peak torque—torque value, determined by the supplier of the drive unit. This torque is provided only for an instant and is normally 200% of the running torque.

Table 1
Gravity Thickener Surface Area Design Criteria

Type of sludge	Influent solids concentration (%)	Expected underflow concentration solids (%)	Mass loading (lb/ft^2/h)[a]
Separate sludges			
Primary (PRI)	2–7	5–10	0.8–1.2
Trickling filter (TF)	1–4	3–6	0.3–0.4
Rotating biological contactor (RBC)	1–3.5	2–5	0.3–0.4
Waste-activated sludge (WAS)			
WAS–air	0.5–1.5	2–3	0.1–0.3
WAS–oxygen	0.5–1.5	2–3	0.1–0.3
WAS–(extended aeration)	0.2–1	2–3	0.2–0.3
Anaerobically-digested sludge from primary digester	8	12	1
Thermally-conditioned sludge			
PRI only	3–6	12–15	1.6–2.1
PRI + WAS	3–6	8–15	1.2–1.8
WAS only	0.5–1.5	6–10	0.9–1.2
Tertiary sludge			
High lime	3–4.5	12–15	1–2.5
Low lime	3–4.5	10–12	0.4–1.25
Alum	–	–	–
Iron	0.5–1.5	3–4	0.1–0.4
Other sludges			
PRI + WAS	0.5–1.5	4–6	0.2–0.6
	2.5–4	4–7	0.3–0.7
PRI + TF	2–6	5–9	0.5–0.8
PRI + RBC	2–6	5–8	0.4–0.7
PRI + iron	2	4	0.25
PRI + low lime	5	7	0.8
PRI + high lime	7.5	12	1
PRI + (WAS + iron)	1.5	3	0.25
PRI + (WAS + alum)	0.2–0.4	4.5–6.5	0.5–0.7
(PRI + iron) + TF	0.4–0.6	6.5–8.5	0.6–0.8
(PRI + iron) + WAS	1.8	3.6	0.25
WAS + TF	0.5–2.5	2–4	0.1–0.3
Anaerobically digested			
PRI + WAS	4	8	0.6
PRI + (WAS + iron)	4	6	0.6

Source: US EPA

[a] Typically, this term is given in lb/ft^2/d. As wasting to the thickener is not always continuous for 24 h, it is a more realistic approach to use lb/ft^2/h.

1 lb/ft^2/h = 4.9 kg/m^2/h.

Table 2
Operating Results at Various Overflow Rates for Gravity Thickeners[a]

Location	Sludge type[b]	Influent solids concentration (%)	Hydraulic loading (gal/ft^2/h)	Mass loading (lb/ft^2/h)	Thickened solids concentration (%)	Overflow suspended solids (mg/L)
Port Huron, MI	P + WAS	0.6	8	0.34	4.7	2500
Sheboygan, WI	P + TF	0.3	18.6	0.46	8.6	400
	P + (TF + A)	0.5	19.0	0.73	7.8	2000
Grand Rapids, MI	WAS	1.2	4.1	0.42	5.6	140
Lakewood, OH	P + (WAS + A)	0.3	25.8	0.6	5.6	1400

Source: US EPA

[a] Values shown are average values only. For example, at Port Huron, MI the hydraulic loading varies between 7 and 9 gal/ft^2/h (300–400 L/m^2/h), the thickened solids in the underflow between 4 and 6% solids; and the suspended solids in the overflow, from 100 to 10,000 mg/L.

[b] P is the primary sludge.
TF is the trickling filter sludge.
WAS is the waste-activated sludge.
1 gal/ft^2/h = 40.8 L/m^2/h.
A is the alum sludge.
1 lb/ft^2/h = 4.9 kg/m^2/h.

Table 3
Typical Uniform Load (*W*) Values

Sludge type	Truss arm W (lb/ft)^a
Primary only (little grit)	30
Primary only (with grit)	40
Primary + lime	40–60
Waste-activated sludge (WAS)	
Air	20
Oxygen	20
Trickling filter	20
Thermal conditioned	80
Primary + WAS	20–30
Primary + trickling filter	20–30

Source: US EPA

[a] Rake arms typically have a tip speed between 10 and 20 ft/min (3–6 m/min)

1 lb/ft = 1.49 kg/m.

3.3.4. *Total Tank Depth*

The total vertical depth of a gravity thickener is based on three considerations:

a. Tank free board.
b. Settling zone (zone of clear liquid and sedimentation zone).
c. Compression and storage zone (thickening zone).

3.3.4.1. Tank Free Board

Tank free board is the vertical distance between tank liquid surface and top of vertical tank wall. It is a function of:

a. Tank diameter.
b. Type of bridge structure: half or full bridge.
c. Type of influent piping arrangement.
d. Whether or not skimming is provided.

Free board will usually be at least 2–3 ft (0.6–0.9 m), although some designers have used free board heights up to 7–10 ft (2–3 m).

3.3.4.2. Settling Zone

This zone encompasses the theoretical zone of clear liquid and sedimentation zone as shown on Fig. 1. Typically 4–6 ft (1.2–1.8 m) is necessary, with the greater depth being for typically difficult sludges, such as WAS or nitrified sludge.

3.3.4.3. Compression and Storage Zone

Sufficient tank volume must be provided so that the solids will be retained for the period of time that is required to thicken the slurry to the required concentration. In addition, sufficient storage is necessary to compensate for fluctuations in solids loading rate. Another consideration is that gas might be produced because of anaerobic conditions or denitrification. Development of these conditions depends on the type of sludge, liquid temperature, and the length of time sludge is kept in the thickener. Plant operating experience has indicated that the total volume in this zone should not exceed 24 h of maximum sludge wasting.

3.3.5. *Floor Slope*

The floor slopes of thickeners are normally more than 2 in. of vertical distance per foot of tank radius (17 cm/m). This is steeper than the floor slopes for standard clarifiers. The steeper slope maximizes the depth of solids over the sludge hopper, allowing the thickest sludge to be removed. The steeper slope also reduces sludge raking problems by allowing gravity to do a greater part of the work in moving the settled solids to the center of the thickener.

3.3.6. *Other Considerations*

3.3.6.1. Lifting Devices

Optimum functioning of a thickener mechanism can be inhibited by heavy accumulation of solids resulting from power outages or inconsistent accumulations of heavy or viscous sludges. Thickeners can be provided with either a manual or an automatic lifting device that will raise the mechanism above these accumulations. This device has not been considered necessary in the majority of municipal wastewater treatment plants except in applications involving very dense sludges (e.g., thermally-conditioned sludge or primary plus lime sludge).

3.3.6.2. Skimmers

Several years ago, it was rare for skimmers to be installed on gravity thickeners. Today it is a common practice to specify skimming and baffling for new plants. The reason for the change is the increased processing of biological sludges and the inherent floating scum layer associated with those sludges.

3.3.6.3. Polymer Addition

Addition of polymer to gravity thickener feed has been practised at several plants (34,35). Results indicate that the addition of polymers improves solids capture but has little or no effect on increasing solids underflow concentration.

3.3.6.4. Thickener Supernatant

Thickener supernatant or overflow is normally returned to either the primary or secondary treatment process. As indicated in Table 2, the strength of the overflow, as measured by total solids, can vary significantly. The liquid treatment system must be sized to handle the strongest recycled load.

3.3.6.5. Pickets

Stirring with pickets in gravity thickeners is thought to help consolidate sludge in the thickening zone (36). However, the support rake mechanism usually can provide sufficient sludge mixing to make special pickets unnecessary.

3.3.6.6. Feed Pump and Piping

The following guidelines are applicable for feed pump and piping:

a. Use positive displacement feed pumps with variable speed drives for variable head conditions and positive feed control.
b. Provide continuous pumping as much as possible.
c. Design piping for operational flexibility.

3.3.6.7. THICKENER UNDERFLOW PUMP AND PIPING

For variable head conditions and typical abrasiveness of many sludges, a positive displacement pump with variable speed drive should be used and its operations should be controlled by some types of solids sensor, for example, either by a sludge blanket level indicator or solids concentration indicator. Pumps should be located directly adjacent to the thickener for the shortest possible suction line. A positive or pressure head should be provided on the suction side of the pump. A minimum of 10 ft (3 m) should be provided for primary sludges and a minimum of 6 ft (2 m) for all other sludges. It is critical to provide adequate cleanouts and flushing connections on both the pressure and suction sides of the pump. Cleanouts should be brought to an elevation more than that of the water surface so that the line may be rodded without emptying the thickener.

3.3.6.8. OTHER DETAILS

For more information and detailed discussions, the reader is referred to the most recent literature reviews on biosolids and sludge management for the years 2000 (37), 2001 (38), and 2003 (7).

4. COST

4.1. Capital Cost

Several recent publications have developed capital cost curves for gravity thickeners (39–41). Probably the most factual is the reference based on actual US EPA bid documents for the years 1973–1977 (40). The US EPA Report on Construction Costs for Municipal Wastewater Treatment Plants indicates that although the plants cost data was rather scattered, a regression analysis of the data indicated that the capital cost could be approximated by a mathematical relationship (40). The reported relationship is shown in Eq. (2) after adjustment to the 2006 USD using the US Army Corps of Engineers Construction Cost Index for Utilities (*see* Appendix).

$$C = 7.85 \times 10^4 Q^{1.10} \tag{2}$$

where C is the capital cost of process in 2006 USD; Q is the plant design wastewater flow in MGD.

The associated cost include cost for excavation, process piping, equipment, concrete, and steel. In addition, cost for administrating and engineering is equal to 0.2264 of capital cost (40):

$$C_{ae} = 0.2264\ \mathrm{C} \tag{3}$$

where C_{ae} is the administration and engineering cost of process.

4.2. Operating and Maintenance Cost

4.2.1. Labor and Management

Figure 3 shows annual man-hour requirements for O/M. As an example, for a gravity thickener surface area of 1000 ft^2 (93 m^2), a designer would include 350 man-hours of O/M in the cost analysis.

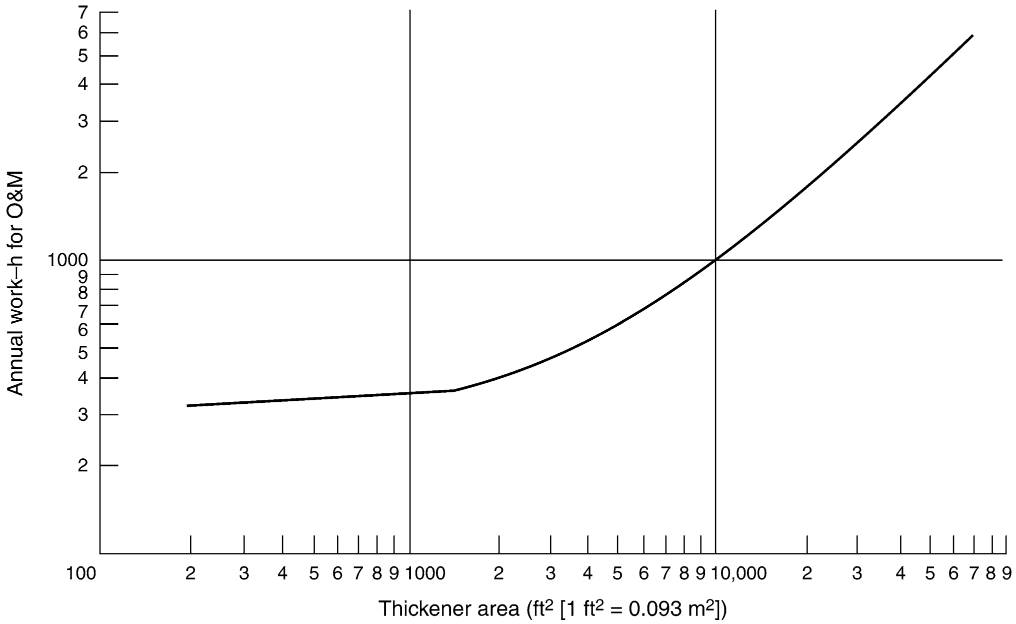

Fig. 3. Annual O&M man-hour requirements for gravity thickeners.

4.2.2. Power Consumption

Figure 4 indicates annual power consumption for a continuously operating gravity thickener as a function of gravity thickener surface area. As an example, for a gravity thickener surface area of 1000 ft^2 (93 m^2), a designer would include a yearly power usage of 4500 kWh (16.2 GJ) in the cost analysis. Figure 4 does not include accessories such as pumps or polymer feed systems.

4.2.3. Maintenance Material Costs

Table 4 is used for estimating circular gravity thickener maintenance material costs as a function of gravity thickener surface area. As an example, for a gravity thickener surface area of 4000 ft^2 (372 m^2), a designer would estimate a yearly materials cost of 2710 USD after adjustment to the 2006 USD using the US Army Corps of Engineers Construction Cost Index for Utilities (Appendix).

5. DESIGN OF THICKENERS

In the design of thickeners, concentration of the underflow and clarification of the overflow must be achieved. Mechanical thickeners (Fig. 5) are designed on the basis of hydraulic surface loading and solid loading. These parameters are normally obtained from laboratory batch settling tests. Procedures for conducting the tests and evaluating the design parameters are well documented in literature. In the absence of laboratory data, Table 1 may be used as a guide for selecting solid loading rates. Typical surface loading rates of 600–800 gpd/ft^2 are recommended for most thickeners. Hydraulic loading rates of less than 400 gpd/ft^2 were reported to produce odor problems. Detention time

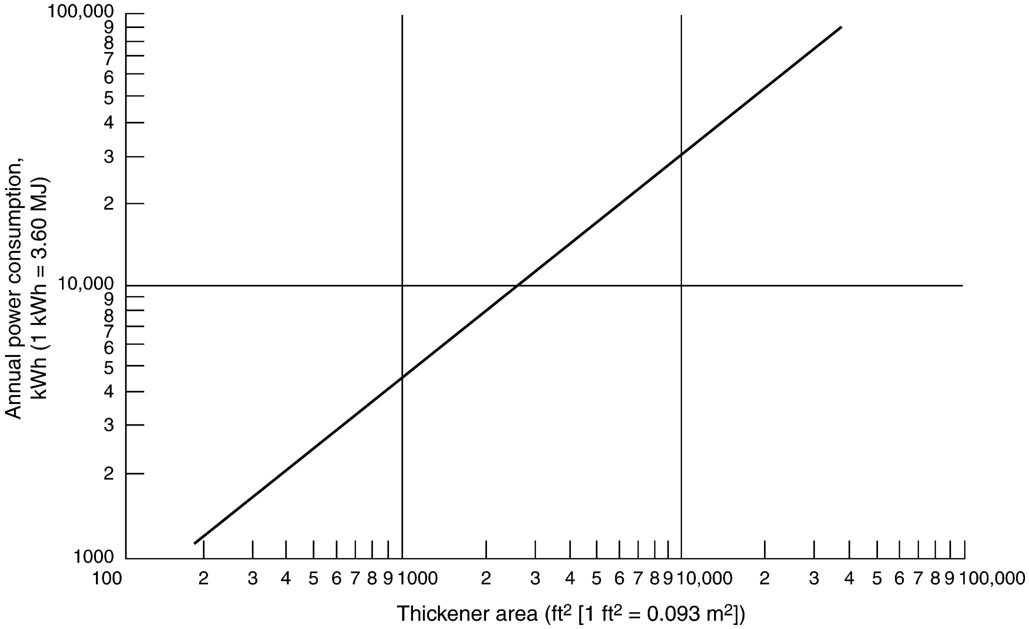

Fig. 4. Annual power consumption for continuously operating gravity thickeners.

Table 4
Maintenance Material Cost for Circular Gravity Thickeners in 2006 USD[a]

Thickener (ft^2)	Area (m^2)	Annual cost (USD)
500	46	630
1000	93	1040
2000	186	1670
4000	372	2710
6000	558	3650
8000	744	4590
10,000	930	5420
20,000	1860	8970
30,000	2790	11,370
40,000	3720	14,080

Source: US EPA

[a] Adjusted to the 2006 USD using the US Army Corps of Engineers Construction Cost Index for Utilities (Appendix).

of the thickener may range between 2 and 4 h. Gravity-thickening is the most common method currently used at wastewater treatment plants for concentrating sludges.

5.1. Input Data

1. Sludge flow (Q) (gpd).
   a. Average daily flow (Q_{avg}) (gpd).
   b. Maxim flow (Q_{max}) (gpd).
   c. Minimum flow (Q_{min}) (gpd).

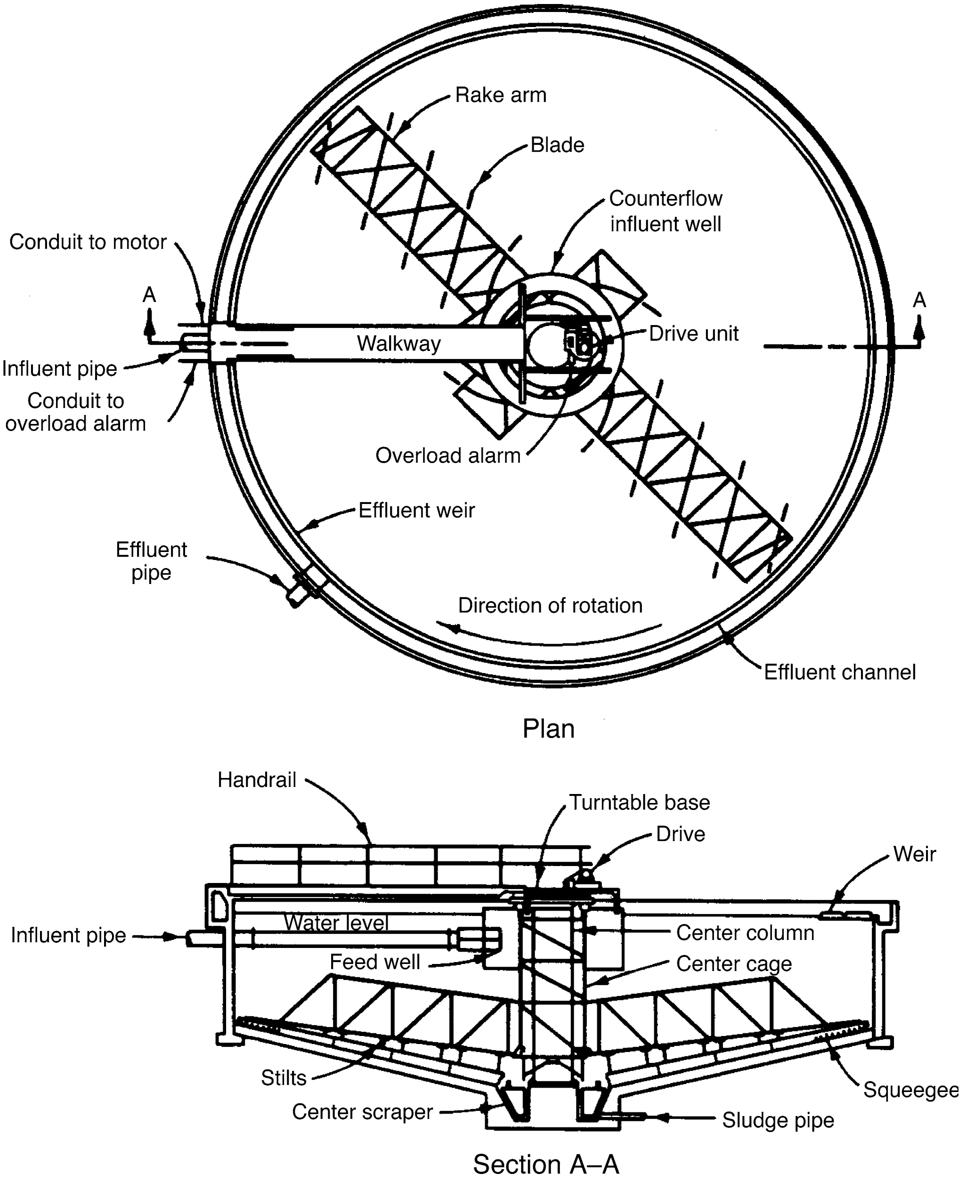

Fig. 5. Mechanical thickener–top and cross-section views.

2. Solid concentration (C_0) (percentage = mg/L × 10^{-4}).
   Note: 1% solids = 0.624 lb/ft^3.

5.2. Design Parameters

a. Desired underflow concentration (C_u) (lb/ft^3 = mg/L × 0.624 × 10^{-4}).
b. Mass loading (ML) (lb/ft^2/d [from settling test]).
c. Hydraulic loading (HL) (gpd/ft^2 [400–800 gpd/ft^2]).
d. Detention time (t) (h [2–6 h]).
e. Number of tanks (N).

5.3. Design Procedure

a. Calculate unit area, using data from settling test (*see* Settling Curve in Fig. 6). Settling of particles from a suspension with high concentration of suspended solids usually involves both zone settling and compression settling in addition to free settling. The compression-settling region is formed under the zone-settling region in a settling column. Settling tests are usually required to determine the sedimentation characteristics of suspensions where zone settling and compression settling occur. Talmadge and Fitch (42) developed a method to determine the area required for a solids handling system from the results of settling tests. This method as outlined by Shammas et al. (18) is given next.

   A settling test is performed with suspension of solids of uniform concentration (C_0) in a settling column of height (H_0). The position of the interface with time is determined and is plotted on a depth time graph (*see* Fig. 6).

   The critical area for a solid handling system is given by the equation

$$A = \frac{Qt_u}{H_0} \tag{4}$$

   where A is the area required for the solids handling thickener (m^2 [ft^2]), Q is the volumetric flow rate into thickener (m^3/s [ft^3/s]), t_u is the time required to attain underflow concentration (c_u) (s), and H_0 is the initial column height of the interface in the settling column (m [ft]).

   In the earlier equation, Q and H_0 are known and t_u is found graphically from the settling curve. First, the point of critical concentration c_c is determined by bisecting the angle formed by extending the tangents to the hindered settling and the compression-settling regions of the settling curve. The bisector cuts the settling curve near the point where compression-settling starts. The critical concentration corresponds to the largest cross-section area required for a solid handling system. The value of t_u can be determined by drawing a vertical line to the time axis from the intersection of the tangent at C_c and the horizontal line drawn at depth H_u· H_u is the depth at which all solids are at the desired underflow concentration (C_u).

$$H_u = \frac{C_0 H_0}{C_u} \tag{5}$$

   where H_u is the the depth at which all solids are at the desired underflow concentration (c_u) (m [ft]), C_u is the underflow concentration; and C_0 is the initial concentration at depth H_0. Knowing the value of t_u, the area required for solids handling system can be obtained from Eq. (4).

b. Calculate mass loading.

$$\mathrm{ML} = \frac{C_0 Q}{A} \tag{6}$$

   Because $Q = V/t_u = H_0 A/t_u$, then:

$$\mathrm{ML} = \frac{C_0 H_0}{t_u} \tag{7}$$

c. If settling data is not available, select mass loading from Table 1.

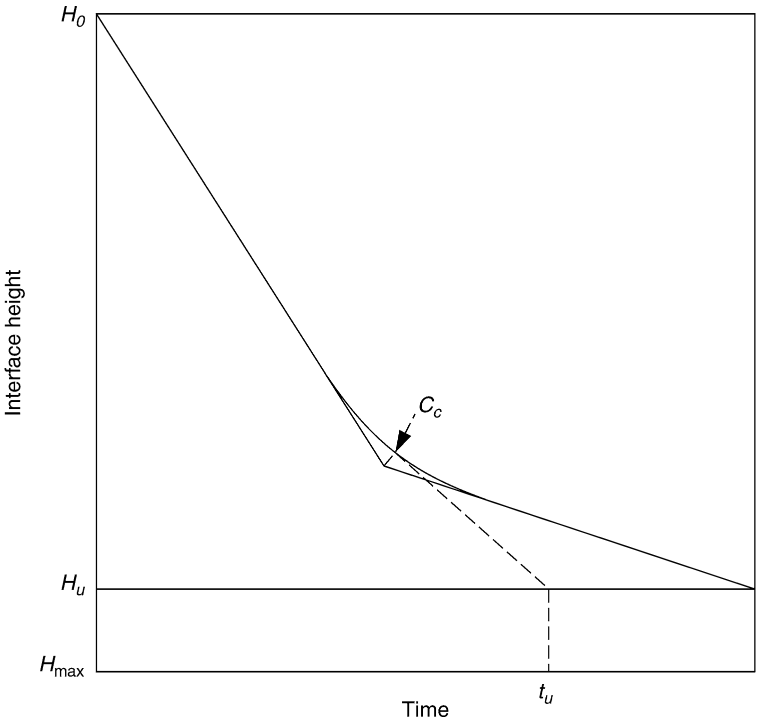

Fig. 6. Graphical determination of t_u, to obtain desired C_u from lab settling curve (18).

d. Calculate total surface area.

$$A = \frac{Q_{avg} \times C_0 \times 0.624}{ML \times 7.48} \tag{8}$$

where A is the total surface area (ft^2); Q_{avg} is the average daily flow (gpd); C_0 is the initial solids concentration (%); and ML is the mass loading (lb/ft^2/d).

e. Check hydraulic loading.

$$HL = \frac{Q_{min}}{A} \quad (>400 \text{ gpd/ft}^2) \tag{9}$$

$$HL = \frac{Q_{max}}{A} \quad (<800 \text{ gpd/ft}^2) \tag{10}$$

where, HL is the hydraulic loading (gpd/ft^2); Q_{min} is the minimum flow (gpd); Q_{max} is the maximum flow (gpd); and A is the surface area (ft^2).

f. Select number of tanks and calculate surface area per tank.

$$A_{tank} = A/N \tag{11}$$

where A_{tank} is the surface area per tank (ft^2); A is the total surface area (ft^2); and N is the number of tanks.

g. Select a detention time (2–6 h) and calculate tank volume.

$$V = (Q_{avg})\,(t)\,(1/24)\,(1/7.48) \tag{12}$$

where V is the tank volume (ft^3); Q_{avg} is the average daily flow (gpd); and t is the detention time (h).

h. Calculate depth.

$$D = \frac{V}{A_{\text{tank}}} \tag{13}$$

where D is the depth (ft); V is the volume (ft^3); and A_{tank} is the surface area per tank (ft^2).

i. Calculate volume of thickened sludge.

$$V_{\text{ts}} = (Q \times C_0 \times \text{recovery}[\%]/100)/C_u \times S_s) \tag{14}$$

where V_{ts} is the volume of thickened sludge per day (gal); Q is the sludge flow (gpd); C_0 is the initial solid concentration (%); C_u is the desired underflow concentration (%); and S_s is the specific gravity of sludge.

5.4. Output Data

a. Average sludge flow (MGD).
b. Initial concentration (%).
c. Thickened concentration (%).
d. Mass loading (lb/ft^2/d).
e. Hydraulic loading (gpd/ft^2).
f. Detention time (h).
g. Number of units.
h. Depth (ft).
i. Volume (ft^3).
j. Surface area per tank (ft^2).
k. Volume of thickened sludge per day (gal).

6. DESIGN EXAMPLE 1

A designer has calculated that it is necessary to thicken a maximum of 2700 lb/d (1225 kg/d) of waste sludge, (dry weight). The sludge consists of 1080 lb (490 kg) of primary at 4% solids and 1620 lb (735 kg) of WAS at 0.8% solids. Waste process from the primary clarifier will be initiated by a time clock and terminated by a sludge density meter when the sludge concentration drops less than a given value. WAS will be pumped from the final clarifier 24 h/d at 17 gpm (1 L/s).

Solution

6.1. Thickener Surface Area

Because this is a new facility and pilot testing is not possible, the designer must utilize Table 1. There are two possible thickening alternatives:

a. The first alternative is thickening of straight WAS with a maximum influent solid concentration of 0.8% solids. At maximum conditions, the designer has selected an average mass loading value of 0.2 lb/ft^2/h (1.47 kg/m^2/h) and a solids concentration of 2% in the underflow.

$$\text{Surface area} = (1620\ \text{lb/d})/(0.2\ \text{lb/ft}^2/\text{h})\ (24\ \text{h/d}) = 337.5\ \text{ft}^2\ (31.4\ \text{m}^2)$$

b. The second alternative is thickening a combination of WAS and primary sludge. The density meter on the primary clarifier will be set to allow the sludge pump to continue as long as the solid concentration is more than or equal to 4% solids. The primary sludge pump will be equipped with a variable speed controller and has a maximum rated pumping capacity of 10 gpm (38 L/min).

On a mass-loading basis, the designer's past experience indicates that surface area required for the combination of primary and WAS is less than that required for WAS alone. However, to assure system reliability, sufficient surface area should be provided to thicken only WAS. With the addition of primary sludge, the expected underflow of solid concentration is 4% (*see* Table 1).

6.2. Hydraulic Loading

The maximum possible hydraulic flow to the gravity thickener would be 17 gpm (1 L/s) of WAS and 10 gpm (0.63 L/s) of primary sludge. The designer is cognizant of the solid recycle problem from the thickener overflow and has selected a value of 6 gal/ft^2/h (250 L/m^2/h) as the maximum overflow rate.

$$\text{Surface area} = \frac{(17+10)\,\text{gpm} \times 60\ \text{min/h})}{6\ \text{gal/ft}^2/\text{h}} = 270\ \text{ft}^2\ \ (25.1\ \text{m}^2)$$

The area required for hydraulic loading is less than that required for mass loading. Hence, the required surface area for thickening is 337.5 ft^2 (31.4 m^2). Because continuous operation of the sludge handling system is essential, two gravity thickeners, each capable of handling the sludge flow, will be provided. The minimum required area for each tank is 337.5 ft^2 (31.4 m^2), which is equivalent to a 20.7 ft (6.2 m) diameter unit:

$$\frac{\pi D^2}{4} = 337.5\ \text{ft}^2$$

hence, $D = 20.7$ ft.

In this size range, equipment manufacturers have standardized on 1 ft (0.3 m) increments; therefore, a 21 ft (6.3 m) diameter, 346 ft^2 (32.2 m^2) units will be specified.

6.3. Torque Requirements

From Table 3, choose a truss arm loading (W) of 30 lb/ft (45 kg/m). From Eq. (1), the running torque required is:

$$T = WR^2 \tag{1}$$

$$T = (30\ \text{lb/ft}) \times (10.5\ \text{ft})^2 = 3307\ \text{ft lb}\ (465\ \text{m kg})$$

The designer will specify a minimum running torque capacity of 3307 ft lb (465 m kg). The other torques will be as follows:

a. Alarm torque = 120% × 3307 = 3970 ft lb (560 m kg).
b. Shut-off torque = 140% × 3307 = 4630 ft lb (650 m kg).
c. Peak torque = 200% × 3307 = 6610 ft lb (930 m kg).

6.4. Tank Depth

Because both the full and the half bridge systems work equally well and the full bridge is less expensive to install, the designer will use a full bridge thickener mechanism that will rest atop the gravity thickener and will have a skimming mechanism attached. In order to accommodate the skimming arm beneath the bridge and allow room to perform maintenance work, the designer selected 2 ft (0.61m) for the freeboard in the thickener. From past experience, the designer selected a typical depth of 5 ft (1.54 m) for the settling zone.

To calculate the depth of the thickening zone, it is assumed that the average solid concentration in the zone would be 1.4% solids and that 1-d storage would be utilized. The following assumptions were made in order to arrive at this percentage:

a. Only WAS would be thickened.
b. The top of the thickening zone would hold 0.8% solids.
c. The bottom of the thickening zone would hold 2% solids

Hence, The average concentration would be equal to:

$$0.8 + 2/2 = 1.4\%$$

Depth of thickening zone = 1620 lb of WAS/(0.014) (8.34) (7.48 gal/ft^3) (346 ft^2) = 5.36 ft (1.61 m).

The total vertical sidewall depth of the gravity thickener is the sum of the freeboard, settling zone, and required thickening zone. In this case total sidewall depth would be:

$$= 2 + 5 + 5.36$$

$$= 12.36 \text{ ft } (3.77 \text{ m})$$

At this time, no allowance has been made for the depth of the cone height of the thickener, which would reduce slightly the vertical sidewall depth of the thickening zone when subtracted from the thickening zone depth.

7. DESIGN EXAMPLE 2

Design a gravity thickener to thicken sludge from a 2 MGD-activated sludge plant. Both primary and activated sludge will be mixed and thickened together. Provide at least two thickeners. Determine the quantity of sludge expected assuming the sludge is undigested. Sludge volume is usually between 7000 and 10,000 gal/MGD of wastewater flow, and solids in the unthickened sludge are between 1800 and 2600 lb/MGD of wastewater flow.

7.1. Quantity of Sludge and Solids

Solution

Quantity of sludge per day assuming 10,000 gal/MGD is 10,000 × 2 = 20,000 gal.

Quantity of solid mass per day assuming 2000 lb of solids/MGD.

$$\text{Weight of solids} = 2000 \times 2 = 4000 \text{ lb.}$$

7.2. Surface Area of Thickeners

Solution

Assume a hydraulic loading, HL of 500 gpd/ft^2

$$\begin{aligned} A &= Q/\text{HL} \\ &= 20{,}000/500 \\ &= 40 \text{ ft}^2 \text{ or } 20 \text{ ft}^2 \text{ for each thickener.} \end{aligned}$$

Assume solids mass loadings, ML of 8 lb/ft^2/d.

$$\begin{aligned} A \text{ required} &= \text{Weight of solids/ML} \\ &= 4000/8 \\ &= 500 \text{ ft}^2 \text{ or } 250 \text{ ft}^2 \text{ per thickener.} \end{aligned}$$

Therefore solids loading govern the design.

7.3. Diameter of Thickeners

Solution

$$\text{Area} = \pi\, D^2/4.$$

$$\text{Area} = 3.14\, D^2/4 = 250.$$

$$\text{Diameter} = 17.8 \text{ ft.}$$

Provide two thickeners of 20 ft diameter each. As the overflow rate for the anticipated volume of sludge is very low some of the wastewater effluents should be recycled to prevent septic conditions.

8. DESIGN EXAMPLE 3

Settling column analyses performed on a waste sludge with an initial solid concentration of 3500 mg/L yielded the settling curve as plotted in Fig. 7. It is desired to thicken this sludge so that the underflow concentration is 14,000 mg/L. The sludge inflow is 0.02 m^3/s. Determine the area required for thickening, the overflow rate and the solid loading. H_0 is 80 cm.

8.1. Height of Sludge at the Required Solids Concentration (C_u)

Solution

From Eq. (5)

$$H_u = \frac{C_0 H_0}{C_u} \tag{5}$$

$$H_u = \frac{3500 \text{ mg}}{\text{L}} \times 80 \text{ cm} \times \frac{\text{L}}{14{,}000 \text{ mg}} = 20 \text{ cm}$$

8.2. Surface Area of Thickener

Solution

Draw tangents to the hindered settling and the compression settling regions of the settling curve, as shown in Fig. 7. Bisect the angle formed by these tangents and obtain the point of critical concentration c_c. Determine the value of t_u by drawing a vertical line to the time axis from the intersection of tangent at c_c and the horizontal line drawn at depth H_u (20 cm). As shown in Fig. 7 the value of t_u is 24 min.

From Eq. (4) the area required for thickening is

$$A = \frac{Q t_u}{H_0} \tag{4}$$

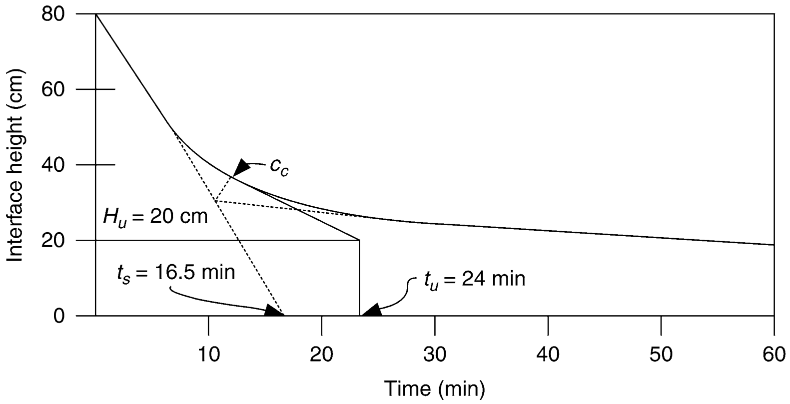

Fig. 7. Settling curve for example 3.

$$= \frac{0.02\ \text{m}^3}{\text{s}} \times 24\ \text{min} \times \frac{60\ \text{s}}{1\ \text{min}} \times \frac{1}{80\ \text{cm}} \times \frac{100\ \text{cm}}{1\ \text{m}}$$
$$= 36\ \text{m}^2$$

While determining the area required to thicken sludge it is essential to ensure that this area is adequate for clarification. The settling velocity, v_s, of the sludge is obtained by dividing the depth of the column by the time, t_s, where the tangent to hinder settling region cuts the time axis.

$$v_s = \frac{80}{16.5 \times 60} = 0.08\ \text{cm/s}$$

The area required for clarification equals

$$A = \frac{Q}{v_s} = \frac{0.02\ \text{m}^3}{\text{s}} \times \frac{\text{s}}{0.08\ \text{cm}} \times \frac{100\ \text{cm}}{1\ \text{m}} = 25\ \text{m}^2$$

which is less than that required for thickening. Therefore area for thickening governs the design.

8.3. Solid Loading

Solution

From Eq. (6) solids mass loading equals:

$$\text{ML} = Q\,C_0/A$$

$$\text{ML} = \frac{0.02\ \text{m}^3}{\text{s}} \times \frac{3500\ \text{mg}}{L} \times \frac{1}{36\ \text{m}^2} \times \frac{86{,}400\ \text{s}}{\text{d}} \times \frac{1000\ L}{\text{m}^3} \times \frac{\text{g}}{1000\ \text{mg}}$$
$$= 168{,}000\ g/m^2/d = 168\ kg/m^2/d \qquad (6)$$

$$\text{ML} = 168/24$$

$$\text{ML} = 7\ \text{kg/m}^2\text{/h}$$

NOMENCLATURE

A	Thickeners area required for solids handling (ft^2 [m^2]).
A_{tank}	Surface area per tank (ft^2 [m^2]).
C	Capital cost (USD).
C_{ae}	Administration and engineering cost (USD).
C_o	Initial solids concentration at height H_o (% [mg/L]).
C_u	Underflow concentration (% [mg/L]).
D	Depth of tank (ft [m]).
D	Diameter of tank (ft [m]).
H_0	Initial column height of the interface in the settling column (ft [m]).
H_u	The depth at which all solids are at the desired underflow concentration (C_u) (ft [m]).
HL	Hydraulic loading [gpd/ft^2 [L/d/m^2]).
Q	Wastewater flow (MGD [ML/d]).
Q	Sludge flow {gpd (ft^3/s [L/d, m^3/s])}.
Q_{avg}	Average daily flow (gpd [L/d]).
$Q_{\max}$	Maximum flow (gpd [L/d]).
$Q_{\min}$	Minimum flow (gpd [L/d]).
ML	Mass loading (lb/ft^2/d [kg/m^2/d]).
N	Number of tanks.
R	Tank radius (ft [m]).
S_s	Specific gravity of sludge.
t	Detention time (h)
t_s	Time as defined in Fig. 7 (s [min]).
t_u	Time required to attain underflow concentration (C_{u}) (s [min]).
T	Torque [ft lb (m kg)].
v_s	Settling velocity at time t_{s} as defined in Fig. 7 (ft/s [cm/s]).
V	Volume (ft^3 [m^3]).
V_{ts}	Volume of thickened sludge per day (gal [L]).
W	Uniform load in Table 3 (lb/ft [kg/m]).

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APPENDIX

United States Yearly Average Cost Index for Utilities[a]

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

[a] Extracted from ref. 43.

3
Flotation Thickening

Lawrence K. Wang, Nazih K. Shammas, William A. Selke, and Donald B. Aulenbach

CONTENTS

1. INTRODUCTION

1.1. Flotation Processes

Flotation is a solid–liquid separation process. Separation is artificially induced by introducing fine gas bubbles (usually air) into the flotation process system. The gas bubbles become attached to the solid particulates, forming a gas–solid aggregate with an overall bulk density less than the density of the liquid; thus, these aggregates rise to the surface of the fluid. Once the solid particles have been floated to the surface, they can be collected by a skimming operation (1–15).

In potable water treatment and wastewater treatment, flotation is used successfully as a clarification process to remove coagulated/flocculated impurities and suspended solids (16–28). In sludge treatment, flotation is used as a thickening process to concentrate various types of organic and chemical sludges (14–16,29–32).

Air flotation systems may be classified as: (a) dispersed air flotation and (b) dissolved air flotation (DAF). In dispersed air flotation, air bubbles are generated by introducing air through a revolving impeller or porous media. This type of flotation system finds some special application in wastewater treatment when wastewater contains surface-active

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agents (9–11) or separate suspended particles on the basis of its surface energy. Another chapter of this handbook entitled, “Dispersed Air Flotation and Electroflotation” presents the dispersed air flotation process in detail (11).

DAF may be subclassified as: (a) pressure flotation and (b) vacuum flotation. Pressure flotation involves air being dissolved in the wastewater under elevated pressures and later released at atmospheric pressure. Vacuum flotation, however, consists of applying a vacuum to wastewater aerated at atmospheric pressure. Dissolved air-pressure flotation, considered herein, is the most commonly used in treatment of potable water, wastewater, and sludge. Since the applications of DAF for water and wastewater treatment are introduced elsewhere (12), this chapter emphasizes only DAF thickening processes.

Other flotation processes, which can be used for sludge thickening include:

a. Electroflotation (11,16).
b. Biological flotation (11,16,33,34).
c. Sequencing batch flotation (17,19,35).

Electroflotation and sequencing batch flotation are discussed elsewhere in this handbook series in detail (11,35). Biological flotation was initially developed by Wang (33), and successfully demonstrated by the Lenox Institute of Water Technology (16). Biological flotation adopts the chemistry of nitrification and denitrification for generation of fine nitrogen and carbon dioxide bubbles in its flotation thickening process reactors for concentration of mainly waste activated sludges (WAS) (16,33,34,36).

1.2. DAF Thickener Components

The principal components of a dissolved air-pressure flotation system (Fig. 1) are a pressurizing pump, air injection facilities, a pressure retention tank, a backpressure-regulating device, usually a throttling valve, and a flotation unit. The primary variables for flotation design are pressure, recycle ratio, feed solid concentration, detention period, air-to-solids ratio, and solids and hydraulic loadings. Optimum design parameters must be obtained from bench scale or pilot plant studies.

1.3. DAF Thickener Advantages and Disadvantages

Since the 1957 installation of the first municipal DAF thickener in the Bay Park Sewage treatment plant, Nassau County, New York, about 300 US municipal installations (over 700 units) have been installed. Although the principal use of the DAF thickener has been to thicken WAS, about 20% of the installations handle other sludge types. Table 1 lists the types of municipal wastewater sludges currently being thickened by DAF thickeners.

The following are the advantages and disadvantages of DAF thickeners compared with other major thickening equipment. The advantages of a DAF thickener are:

a. It provides better solids–liquid separation than a gravity thickener.
b. For many sludges, it yields a higher concentration of solids than a gravity thickener.
c. It requires less area than a gravity thickener.
d. It offers excellent sludge equalization control.
e. It has less chance of odor problems than a gravity thickener.
f. It can remove grit from a sludge processing system.
g. It removes grease.

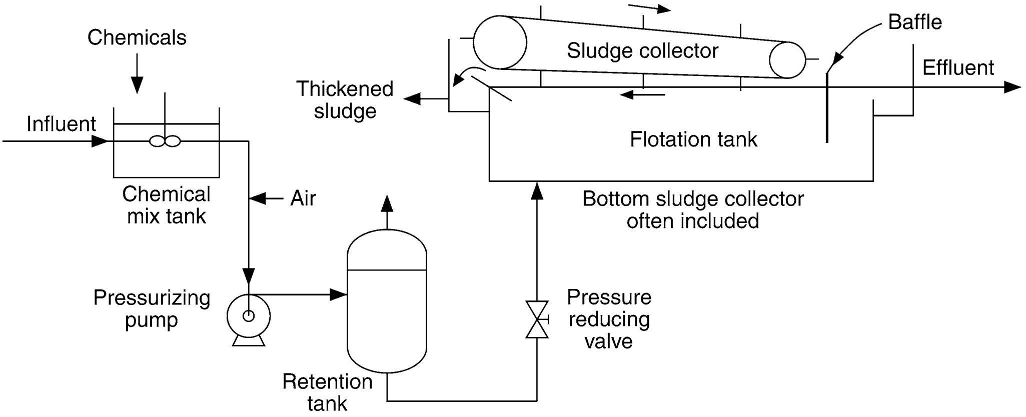

Fig. 1. Rectangular DAF thickener flow diagram (*Source:* US EPA).

Table 1
Design Criteria of DAF Thickening Process Based on the Types of Sludges

Sludge type	Feed solids concentration (%)	Typical loading rate without polymer ($lb/ft^2/d$)	Typical loading rate with polymer ($lb/ft^2/d$)	Float solids concentration (%)
Primary + WAS	2	20	60	5.5
Primary + (WAS + $FeCl_3$)	1.5	15	45	3.5
(Primary + $FeCl_3$) + WAS	1.8	15	45	4
WAS	1	10	30	3
WAS + $FeCl_3$	1	10	30	2.5
Digested primary + WAS	4	20	60	10
Digested primary + (WAS + $FeCl_3$)	4	15	45	8
Tertiary (alum)	1	8	24	2

Source: US EPA.
WAS = waste activated sludge.

The disadvantages of a DAF thickener are:

a. Its operating cost is higher than for a gravity thickener with equal flow capacity.
b. DAF-thickened sludge concentration is less than in a centrifuge.
c. It requires more area than a centrifuge.
d. It has very little sludge storage capacity.

2. DAF THICKENER PROCESS DESCRIPTION

In a DAF thickening process, air is added at pressures in excess of atmospheric pressure (30–70 psig) either to the incoming sludge stream or to a separate liquid stream. When pressure is reduced and turbulence is created, air in excess of that required for saturation at atmospheric pressure, converts the solution into very small bubbles of 50–100 μm in diameter. The bubbles adhere to the suspended particles or become enmeshed in the solids matrix. Since the average density of the solids–air aggregate is less than that of water, the agglomerate floats to the surface. The floated solids build to a depth of several inches at the water surface. Water drains from the float and affects solids concentration. Skimmers continuously remove the float. Good solids flotation occurs with a solids–air aggregate specific gravity of 0.6–0.7. Polyelectrolytes are frequently used as flotation aids, to enhance performance and create a thicker sludge blanket (12). A description of the DAF thickener process in general has been represented in Fig. 1.

DAF thickeners can be utilized either to thicken wastewater solids prior to dewatering or stabilization or to thicken aerobically digested or other solids prior to disposal or dewatering. In terms of reactor shape, materials and operation, DAF thickeners can be rectangular or circular, constructed of concrete or steel, and can operate in the full, partial, or recycle pressurization modes (15,15,20).

2.1. Full Pressurization DAF-Thickening System

There are three ways in which a DAF system can be operated. The first method is called "full or total pressurization." With this design, the entire sludge flow is pumped through the pressure retention tank, where the sludge is saturated with air and then passed through a pressure reduction valve before entering the flotation chamber. As in the other two modes, a distribution device is used to dissipate inlet energy and thus prevents turbulence and limits short-circuiting. The primary advantage of pressurizing the total flow is that it minimizes the size of the flotation chamber, a significant part of the capital cost. However, the advantage of a smaller chamber may be partially offset by the cost of a higher head feed pump, larger pressure vessel, and more expensive operation. Operational problems may result from floc shearing and clogging when sludge is passed through the pressure-regulating valve (20).

2.2. Partial Pressurization DAF-Thickening System

The second method of operation is called "partial pressurization." With this design only part of the sludge flow is pumped through the pressure retention tank. After pressurization the unpressurized and pressurized streams are combined and mixed before they enter the flotation chamber. In this arrangement the pressurizing pump and pressure vessel are smaller and the process is not as susceptible to flow variations, as is total pressurization; this is the case when the necessary pump controls are included in the design. The size of the flotation chamber would be the same as that for a total pressurization system.

2.3. Recycle Pressurization DAF-Thickening System

The third method is called "recycle pressurization." Here, a portion of the clarified liquor (subnatant) or an alternate source containing relatively little suspended matter is pressurized. Once saturated with air, it is combined and mixed with the unthickened sludge before it is released into the flotation chamber.

The major advantage of this system over the total and partial pressurization system is that it minimizes high shear conditions, an important parameter when dealing with flocculent-type sludges. Another advantage arises when wastewater sludge streams containing stringy materials are thickened. The recycle pressurization system eliminates clogging problems with the pressurization pump, retention tank, and pressure release valve.

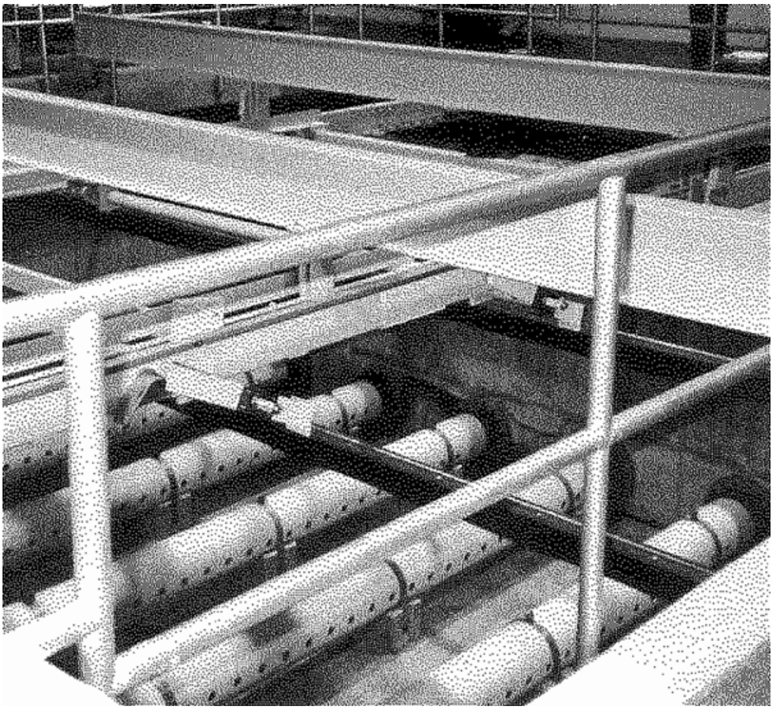

Fig. 2. Rectangular DAF thickener picture (Dongshin, Seoul, Korea).

For the aforementioned reasons, recycle pressurization systems are the most commonly used units in the United States. Figures 1 and 2 show a typical rectangular DAF tank installation. In this system, the pressure retention tank may be either unpacked or packed (meaning that the tank is filled with a packing material to create turbulence). The use of either is dependent principally on the source of the pressurized recycle flow.

The pressurized recycle flow can be obtained either from the subnatant stream or, typically, from the secondary effluent. The advantages of using secondary effluent are that it results in a much cleaner stream (low suspended solids and low grease content) and allows the use of a packed pressure retention tank. A packed tank is smaller than an unpacked tank, has lower associated capital cost, and provides for a more efficient saturation of the liquid stream. In this case, less air is required to achieve the same level of liquid saturation as compared with an unpacked tank and power requirements are lower. However, packed tanks may eventually require cleaning, and the use of secondary plant effluent will significantly increase the flow through the secondary treatment system, thereby increasing pumping costs and possibly affecting the performance of the secondary clarifier.

3. PROCESS APPLICATIONS AND LIMITATIONS

DAF is the most common form of flotation thickening in use in the United States. The process has been used for many years to thicken WAS and to a lesser degree to thicken combined sludges. DAF also has widespread industrial wastewater applications and domestic potable water applications (38). However, only the thickening applications are discussed in this chapter.

3.1. Sludge Thickening Applications

The use of DAF thickening processes is limited primarily to thickening of biological sludges prior to dewatering or digestion at wastewater treatment plants. Used in this way, the efficiency of the subsequent dewatering units can be increased and the volume of supernatant from the subsequent digestion units can be decreased. Existing air flotation thickening units can be upgraded by the optimization of process variables, and by the utilization of polyelectrolytes. Air flotation thickening is best applied to WAS. With this process, it is possible to thicken the sludge to 6% solids, while the maximum concentration attainable by gravity thickening without chemical addition is 2–3% solids. The DAF process can also be applied to mixtures of primary and WAS. DAF also maintains the sludge in aerobic condition and has better potential to capture solids than gravity thickening. There is some evidence that activated sludges from pure oxygen systems are more amenable to flotation thickening than sludges from conventional systems. Recently Krofta and Wang (29–32) have successfully used DAF thickening process for concentration of chemical alum sludges generated at water treatment plants.

3.2. DAF Thickening Process Limitations

So far the DAF thickening process is limited to concentration of WAS at wastewater treatment plants and alum sludge at water treatment plants. The variability of sludge characteristics requires that some pilot work be done before the designing of a DAF system.

4. PROCESS DESIGN CONSIDERATIONS

4.1. Rectangular or Circular Shape

Figures 1 and 2 introduce rectangular DAF thickeners, while Figs. 3 and 4 show circular DAF thickeners, respectively. Both rectangular DAF thickeners and circular DAF thickeners are equally effective for float removal. Rectangular DAF thickener's skimmers can easily be closely spaced; secondly, they can be designed to skim the entire surface. Because of the sidewalls, float does not easily move around the end of the skimmers. Bottom sludge flights are usually driven by a separate unit and, hence, can be operated independently of the skimmer flights. Water level in the tank can be changed readily by adjusting the end weir. This allows changing the depth of water and flight submergence to accommodate changes in float weight and displacement, which affect the ability to remove this material from the unit.

The main advantage of circular units is their lower cost in terms of both structural concrete and mechanical equipment. For example, two 60-ft (18 m) diameter circular units are the equivalent of three 20-ft by 90-ft (6 m by 27 m) rectangular units. The rectangular units require approx 11% more structural concrete, as well as more drives and controls which increase maintenance requirements.

4.2. Concrete or Steel Construction

Steel tanks come completely assembled and only require a concrete foundation pad and piping and wiring hookups. Although equipment purchase price is much higher for steel tanks, considerable field labor and expensive equipment installation are eliminated (15,16,29–32). Structural and shipping problems limit steel DAF units to the smaller sizes, such as 450 ft^2 (40.5 m^2) or less for rectangular units, and 100 ft^2 (9 m^2) for circular units.

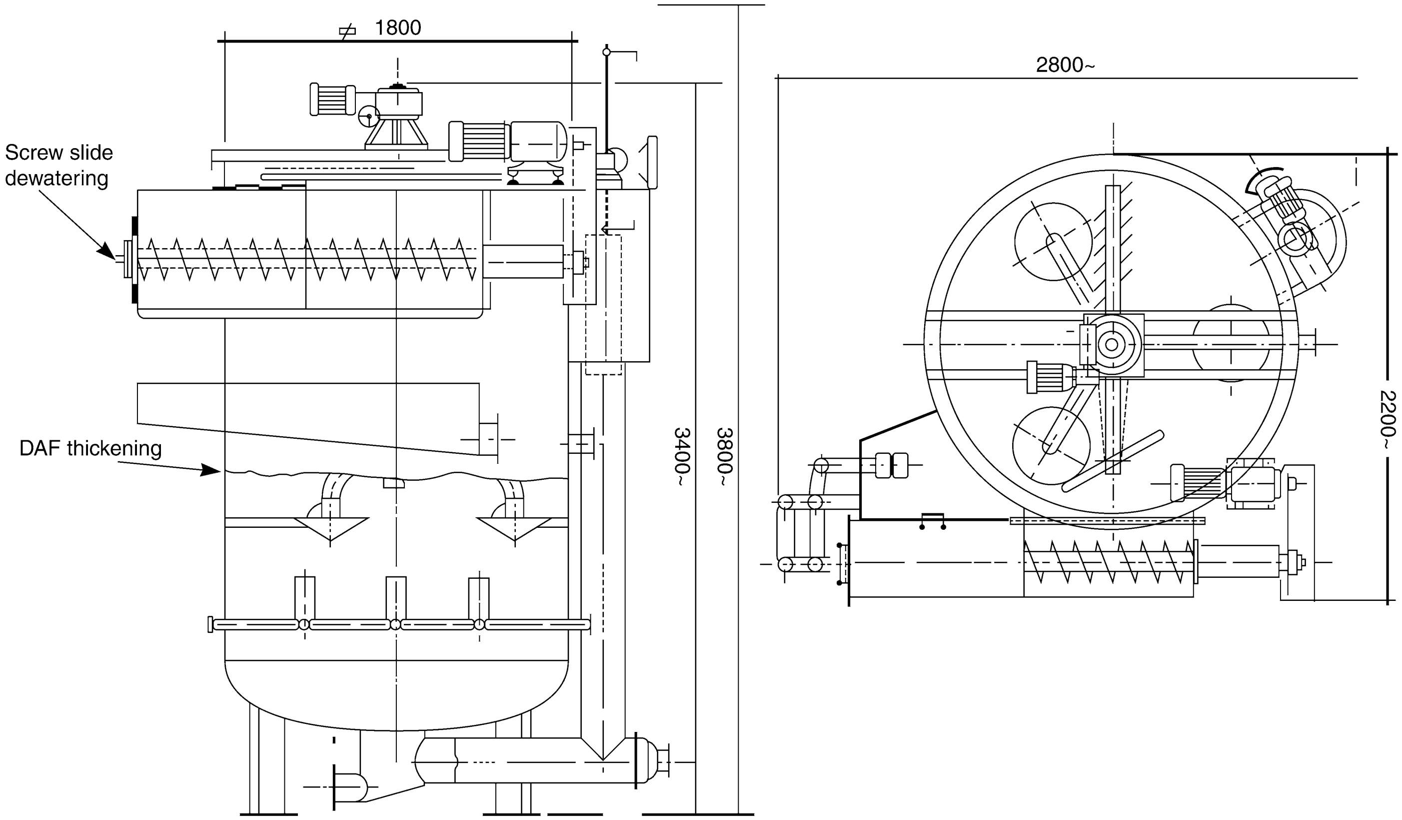

Fig. 3. Circular DAF thickener side and top views (*Source:* Lenox Institute of Water Technology, Lenox, MA).

Fig. 4. Circular DAF thickener picture (Krofta Engineering Corporation, Lenox, MA).

For a large installation requiring multiple tanks or large tanks, concrete tanks are more economical. Once the largest DAF plant (37.5 MGD peak capacity), Pittsfield, MA utilizes concrete flotation tanks (21,28).

4.3. Pilot-Scale or Bench-Scale Experiments

If sludge is available, the designer should, at least perform bench-scale testing. If money is available, consideration should be given to renting a pilot DAF thickener and conducting 4–6 wk test program to evaluate the effects of such parameters as recycle ratio, air-to-solids ratio, solids and hydraulic loading, and polymer type and dosage. If sludge is not available, then a detailed review must be made of experiences at installations where a similar type of sludge is being thickened by DAF thickeners (20).

4.4. Influent Feed Characteristics

The first step in designing a DAF thickener is to evaluate the characteristics of the feed stream. The designer must evaluate the type of sludge(s) to be thickened and the approximate quantities of each under various plant loadings and modes of operation. If WAS is to be thickened, the expected range of the sludge-age must be determined since sludge-age can significantly affect DAF thickening performance. Information is needed about the source of waste sludge and the range of solids concentrations that can be expected. Also, there should be an evaluation of any characteristic of the feed stream that may affect air solubility, for example, concentration of dissolved salts, and range of liquid temperatures.

4.5. Thickener Surface Area

To calculate the effective surface area of a DAF thickener, a designer must know the net solids load, solids surface loading rate, and hydraulic surface loading rate.

4.5.1. Net Solids Load

As DAF thickener is not entirely efficient, more sludge must be pumped into the thickener than the actual amount removed. The actual amount removed is the net solids load. From a design standpoint, the net load is the amount of solids that must be removed from the liquid processing train each day. This value divided by the appropriate solids loading rate gives the required effective surface area. The gross solids load is calculated by dividing the net load by the expected solids capture efficiency of the system. The gross solids load is important in sizing system hydraulic piping.

4.5.2. Solids Loading Rate

The allowable solids loading rate is related to the minimum solids flux that will occur within the range of sludge concentrations found in the thickener. This flux is a function of the type of sludge processed, the float concentration desired, and polymer used. Pounds of dry solids per square foot per day or pounds of dry solids per square foot per hour are the units used to express this rate (16,18).

The effect of sludge type on the solids loading rate is shown in Table 2. The loading rates indicated will normally result in a minimum of 4% solids concentration in the float. Actual operating data is listed in Table 3. In general, increasing the solids loading rate decreases the float concentration. Figure 5 illustrates this phenomenon with polymer addition. The addition of polyelectrolyte will usually increase the allowable solids loading rate.

4.5.3. Hydraulic Loading

The hydraulic loading rate for a DAF thickener is normally expressed as gallons per minute per square foot (gpm/ft^2). When like units are cancelled, the hydraulic loading rate becomes a velocity equivalent to what the average downward velocity of water would be if it were imagined to flow downward through the tank. The maximum hydraulic rate must always be less than the minimum rise rate of the sludge/air particles to ensure that all the particles will reach the sludge float before the particle reaches the effluent end of the tank.

Reported values for hydraulic loading rates range from 0.79 to 4.0 gpm/ft^2 (0.54 to 2.7 L/s/m^2). This wide range probably indicates a lack of understanding of the term. In some cases, the hydraulic loading refers simply to the influent sludge flow, while in others the recycle flow is included. In most sources, no definition of the term was given. Table 4 indicates the hydraulic loading rates found in the literature.

Table 2
Typical DAF Thickener Solids Loading Rates Necessary to Produce a Minimum 4% Solids Concentration

Type of sludge	Solids loading rate (lb/ft^2/h) No chemical addition	Solids loading rate (lb/ft^2/h) Optimum chemical addition
Primary only	0.83–1.25	Up to 2.5
Waste activated sludge (WAS)		
Air	0.42	Up to 2
Oxygen	0.6–0.8	Up to 2.2
Trickling filter	0.6–0.8	Up to 2
Primary + WAS (air)	0.6–1.25	Up to 2
Primary + trickling filter	0.83–1.25	Up to 2.5

Source: US EPA.
1 lb/ft^2/h = 4.9 kg/m^2/h.

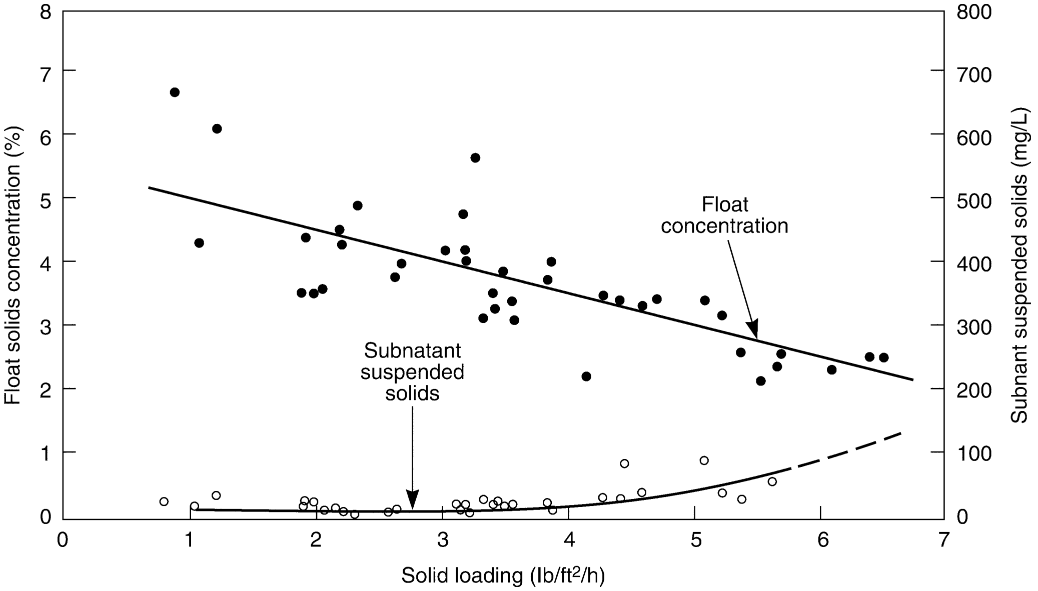

Fig. 5. Float concentration and subnatant suspended solids vs solids loading of a waste activated sludge—with polymers (*Source:* US EPA). 1 lb/ft^2/h = 4.9 kg/m^2/h.

As the total flow through the thickener affects the particles, the hydraulic loading rate should be based on the total flow (influent plus recycle). Extensive research on WAS has resulted in the conclusion that a peak rate of 2.5 gpm/ft^2 (1.7 L/s/m^2) should be employed. This value is based on use of polymers. When polymers are not used, this value is expected to be lower, but no design criterion has been suggested at this time.

Table 3
Field Operation Results from Rectangular DAF Thickeners

Installation	Sludge type[a]	Solids loading rate (lb/ft^2/h)	Feed solids concentration (mg/L)	Polymer dosage (lb/dry t solids)	Float solids concentration (%)	Subnatant suspended solids (mg/L)
Eugene, OR	P + TF	1.25	5000	0	4.5–5	500
Springdale, AR	P + TF	2.5	20,000	7	6.5	200
Athol, MA	A	3.2	8000	2	4	50
Westgate Fairfax, VA	A[b]	7	14,000	1–4	7.3	20
Warren, MI	A[c]		11,000	40	5	200
Frankenmuth, MI	A[c]	0.58	5000	0	3	750
	A[c]		8000	26	3.5–5.5	90
Cinnaminso, NJ	A	2	5000	5	4	250
San Jose, CA	P + A[d]	1.9	23,000	0	7.1	
	P + A[e]	1.6	17,000	0	5.3	
Boise, ID	A	1	4600	0	4	
	A	1.17	5000	3	3.8	500
	A	1.13	5000	6	4	500
Levittown, PA	A	0.54	8000	0	6.5	
	P + A	1	6400	0	8.6	
Xenia, OH	A		4000	30	2.5–3	100
Indianapolis, IN	P + A		10,000	30	3.5–4.2	100–1000
Columbus, OH (Jackson Pike)	A		6000	0	3.2	800
Wayne County, MI	A	0.83	4500	0	4.6	
Dalton, GA	P + A	0.75	12,900	0	6.1	
Middletown, NJ	A	2	10,000	5–6	4	500

Source: US EPA.

[a] P, primary sludge; A, waste-activated sludge; TF, trickling filter sludge.

[b] Oxygen plant.

[c] Considerable brewery waste.

[d] Noncanning season.

[e] Canning season.

1 lb/ft^2/h = 4.9 kg/m^2/h.

1 lb/t = 0.5 kg/T.

1 t = 1 ton (English ton); 1 T = 1 tonne (metric ton) = 1.1025 t.

4.6. Air-to-Solids Ratio

Another design parameter to be considered in DAF thickening is that of the air-to-solids (*A*/*S*) weight ratio, i.e., lb air/lb solids (kg air/kg solids). Theoretically, the quantity of air required to achieve satisfactory flotation is directly proportional to the

Table 4
Reported DAF Thickener Hydraulic Loading Rates (gpm/ft²)[a]

Influent only	Influent plus recycle
–	1.5–2.5
–	2.5
–	1.0–4.0
–	0.79
–	1.25–1.5
0.9	3.0

Source: US EPA.

[a] All values reported are associated with polymer usage.

1 gpm/ft^2 = 40.8 L/min/m^2.

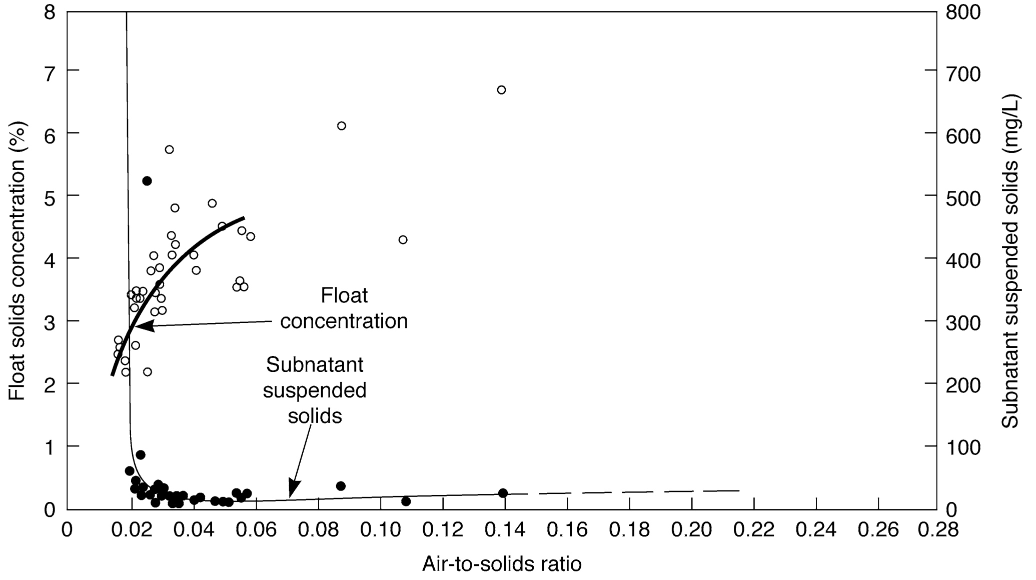

Fig. 6. Float concentration and subnatant suspended solids vs air-to-solids ratio with polymer for waste activated sludge (*Source:* US EPA).

quantity of solids entering the thickener (defined as gross solids load in the previous section). For domestic wastewater sludges, reported ratios range from 0.01 to 0.4, with most systems operating at a value under 0.1 (14,16,18). The appropriate *A*/*S* ratio for a particular application is a function of the characteristics of the sludge, principally, the sludge volume index, the pressurization system's air dissolving efficiency, and the distribution of the gas–liquid mixture into the thickening tank. Figure 6 shows the effects of *A*/*S* of float concentration and subnatant suspended solids, with polymer addition.

4.7. Polymer Usage

Polymers have a marked effect on DAF thickener performance, and therefore a designer must be careful to differentiate between performance with and without polymer use. Polyelectrolytes may improve flotation by substantially increasing the size of the particles present in the waste. The particles in a given waste may not be amenable to the flotation process because their small size will not allow adequate air bubble attachment. The surface properties of the solids may have to be altered before effective flotation can occur. Sludge particles can be surrounded by electrically charged layers that keep these particles dispersed in the liquid phase. Polyelectrolytes can neutralize the charge, causing the particles to coagulate so that air bubbles can attach to them for effective flotation. Thus, with use of polymers, the following operating advantages may occur: the size of the DAF thickener may be reduced; solids capture may be improved, thus reducing the amount of solids recycled back to the liquid handling system; and an existing, overloaded facility in which polymers are not being utilized may be upgraded. They also act as a surfactant, thus allowing better attachment of air bubbles.

The major disadvantage of polymers is cost (polymer cost, operation, and maintenance of polymer feed equipment), when calculated over the useful lifetime of the plant. In addition, the actual amount required is very difficult to determine until flotation studies can be run on the actual installation. If polymers are to be used, it is best to design conservatively, so that the possibility of the exceptionally high polymer demand needed to keep marginal operation at capacity is avoided. Table 3 lists current operating results of plants with and without polymer addition.

4.8. Pressurization System

The air dissolution equipment, which consists of the pressurization pump, air dissolution tank, and other mechanical equipments, is the heart of a DAF thickener system. In sizing a pressurization system, the designer must decide on an operating pressure and a quantity of pressurized flow and must be aware of factors affecting the performance of the system.

4.9. Operating Pressure

Most commercially available pressurization systems operate at 40–80 psig (276–522 kN/m^2). For a given A/S ratio, the air required to float the sludge can be obtained by increasing the operating pressure of the system to dissolve more air, or holding a lower operating pressure and increasing the volume of pressurized flow (4).

The higher the operating pressure of a flotation thickener system, the lower the rise rate of the sludge. The reason for a higher rise rate at 40 psig (276 kN/m^2) than at 60 or 80 psig (414 or 552 kN/m^2) is that the optimum bubble size is predominant at this lower operating pressure. Attempting to raise the A/S ratio by increasing the operating pressure is detrimental to the thickening process. These findings are important in that it will be in the user's best interest to operate at the lowest pressure possible. The requirement for higher head pumps, larger air compressors, and higher pressure rated retention tanks raises the initial cost of the process as well as operating costs.

4.10. Quantity of Pressurized Flow

For a DAF thickener to work effectively, the proper amount of air must be present for each pound of solids to be handled (A/S ratio). The design pressurized flow should be based on the maximum gross solids load that the DAF thickener is designed to receive. For multiple units, each basin should have its own independent pressurization system. This is especially important if the thickening system is designed to operate over a wide range of influent solid concentrations and flows (38).

4.11. Number of Units

The number of DAF thickeners to be provided at a facility depends on the following factors:

a. The availability and configuration of available land.
b. The operating cycle that will be used, for example, 7 d/wk, 24 h/d, 5 d/wk, 8 h/d, and so on.
c. Seasonal variability; for example, the operation of a food processor 6 mo of the year, the waste flow which will go to the municipal facility.
d. The variance in average-to-peak hourly solids load that can be expected on a day-to-day basis.

Since the storage capacity of a DAF system is limited and the averaging of short-term variation is limited, design for the peak hourly waste sludge production is necessary. In addition, provision must be made to handle the sludge flow if a unit must be taken out of service. In addition to the system design considerations previously discussed, consideration must be given to feed sludge line sizing, thickened sludge removal, bottom draw-off piping, subnatant piping, pressurized flow piping, and controls. Each of these items is briefly discussed next.

4.12. Feed Sludge Line

Feed sludge flow rate must be controlled to stay within allowable limits. This requires a flow meter that accurately measures a high solids stream and piping large enough to handle maximum flow.

4.13. Thickened Sludge Removal

The surface skimmer brings the thickened sludge over the dewatering beach and deposits it in a sludge hopper. The thickened sludge must then be pumped to the next phase of the solids handling system. In pump selection, it is important to remember that air has been entrained in this sludge by the flotation thickening process. Pumps that can air lock should not be used; positive displacement pumps are common in this application. For pipe sizing and final pump selection, consider that the thickened sludge can reach concentration in the range of 10%.

4.14. Bottom Sludge Draw Off, Subnatant Line, Pressurized Flow Piping, and Controls

In a rectangular DAF tank, the bottom collector moves the settled solids to the influent end of the basin. Here it is deposited into either multiple hoppers or a cross-screw conveyor that delivers it to a hopper. The bottom collector in a circular DAF tank delivers the settled solids directly to a hopper in the center of the tank. Once the solids are in the hopper, they must be removed from the tank. Depending on where this flow goes, it can be handled by either gravity or pumps.

One major consideration that applies to either removal system, but particularly to gravity removal, is the static head available. Since the draw-off point is at the bottom of the flotation basin, the entire depth of the liquid in the basin must be considered as available static head. Although fine control is not required, this head must be dissipated in order to restrict the flow. A positive displacement pump with variable speed drive will assure control of bottom sludge withdrawal. This draw-off is at the lowest point in the basin and therefore, could also be used as a basin drain. If a tee and drain valve is installed on this line at the outside of the tank wall, draining can take place. The line from the drain valve can go to the plant's drain system.

The subnatant line pipe sizing should be such that it can handle the maximum total flow (influent plus recycle) without any appreciable head loss.

The pressurized flow piping system must be properly designed. Because of the high-pressure requirements of this flow, the pressurization liquor is usually delivered to the pressure tank by a high-speed, closed impeller centrifugal pump. Piping must be sized to handle the maximum liquid throughput rate of the pressure tank selected. The controls for a DAF thickener are dependent on the system, the degree of automation required, and the equipment manufacturer's design. They usually include, at a minimum, a pressure controller for the pressure vessel and flow meters for the feed and thickened sludge flows.

5. PROCESS PERFORMANCE

5.1. Performance Data

A summary of data from various air flotation units indicates that solids recovery ranges from 83 to 99% at solids loading rates of 7–48 lb/ft^2/d.

A summary of operating data from 14 sewage treatment plants with DAF thickeners is as follows:

a. Influent suspended solids = 3000–20,000 mg/L (median 7300 mg/L).
b. Subnatant suspended solids = 31–460 mg/L (median 144 mg/L).
c. Suspended solids removal = 94–99% (median 98.7%).
d. Float solids = 2.8–12.4% (median 5%).
e. Loading = 1.3–7.7 lb/h/ft^2 (median 3.1 lb/h/ft^2).
f. Sludge flow = 0.4–1.8 gal/min/ft^2 (median 1 gal/min/ft^2).

Flotation aids (generally polyelectrolytes) are usually used to enhance DAF thickening process performance. Subnatant (DAF effluent) is good quality and contains approx 150 mg/L TSS. In general subnatant is returned to the mainstream of the wastewater treatment plant for reprocessing.

5.2. Factors Affecting Performance

The designer should be aware of two physical factors, air saturation and turbulence, which can affect the performance of the pressurizing system.

5.2.1. Air Saturation

The basic mechanism that makes flotation possible is the increase in the amount of gas dissolved when pressure is increased (2). The relationship between pressure and quantity dissolved is shown in Henry's Law, which states that if no reaction prevails between the gas and liquid phases, the solubility of the gas is directly proportional to the absolute pressure of the gas at equilibrium with the liquid at constant temperature. In practice, the actual amount of air dissolved for a given air input depends on the efficiency of the pressurization device, liquid temperature and concentration of solutes in the liquid stream being pressurized.

Normally a pressure retention tank is used to maximize the air–water interface for efficient air transfer in the shortest detention time. Depending on tank design (packed tank, unpacked tank, tanks with mechanical mixers, and so on), efficiencies can range from as low as 50 to 90%. It is current design practice in the United States to specify a minimum of 85–90% efficiency.

The equilibrium concentration of a gas in a liquid is inversely related to the temperature of the liquid phase. The temperature effect is substantial. For example, the saturation of air in water at 140°F (60°C) is about one half that of the saturation of air in water at 66°F (18.8°C) at 1 atm. The presence of salts such as chloride will normally decrease the air solubility at a given temperature and pressure. The effect of salt concentration on air dissolving efficiency is best evaluated by conducting bench-scale treatability tests or a pilot unit test program.

5.2.2. Turbulence

The proper amount of turbulence must be present at the point of pressure reduction to cause bubble formation. Without the necessary turbulence, the rate at which air bubbles form is slow and may occur too late in the process. Excessive turbulence can result in increased bubble agglomeration and floc shear. Under this condition, the majority of bubbles formed will be considerably larger than the 50–100 μm needed for effective flotation.

6. PROCESS COST AND OPERATION CONSIDERATIONS

6.1. Capital Cost

Several publications have developed capital cost curves for DAF thickeners. The most factual is the reference based on actual US EPA bid documents. Although the data was scattered, a regression analysis indicated the capital cost could be approximated by Eq. (1), which has been adjusted to 2006 USD using the US Army Corps of Engineers Construction Cost Index for Utilities (*see* Appendix):

$$C = 7.17 \times 10{,}000\ Q^{1.14} \tag{1}$$

where C is the capital cost of DAF thickener process in 2006 (USD) and Q is the plant design flow in MGD.

The associated costs include excavation, process piping, equipment, concrete and steel. In addition, such costs as those for administrating and engineering are equal to 0.2264 times Eq. (1).

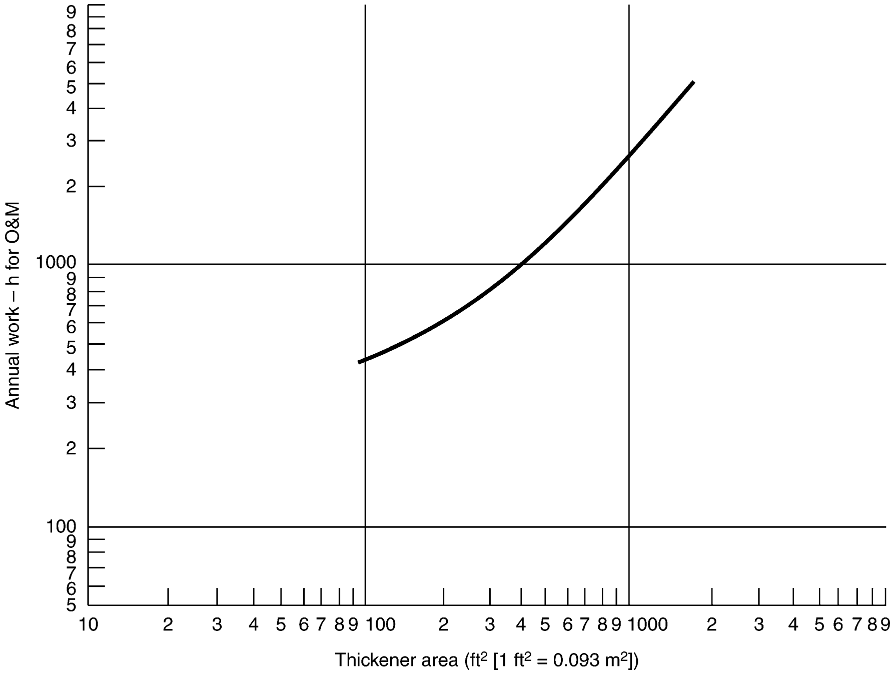

Fig. 7. Annual O&M man-hour requirements for DAF thickeners (*Source:* US EPA).

6.2. Operating and Maintenance Costs

Figure 7 indicates annual man-hour requirements for operations and maintenance. As an example, for a DAF thickener surface area of 1000 ft^2 (93 m^2) a designer would include 2700 man-hours of operation and maintenance in the cost analysis. Figure 8 shows annual power consumption for a continuously operating DAF thickener as a function of DAF thickener surface area. As an example, for a DAF thickener surface area of 1000 ft^2 (93 m^2), a designer would include a yearly power usage of 720,000 kWh (2592 GJ) in the cost analysis. Figure 8 does not include accessories such as pumps or polymer feed systems. Details about the economic sensitivity of the DAF process with respect to the operational variables can be found in the literature (10).

7. PROCESS RELIABILITY AND ENVIRONMENTAL IMPACT

7.1. Reliability

DAF systems are reliable from a mechanical standpoint. Variations in sludge characteristics can affect process (treatment) reliability, and may require operator attention.

7.2. Environmental Impact

The DAF thickener requires less land than gravity thickeners. A subnatant stream is returned to the head of the treatment plant, although it should be compatible with other

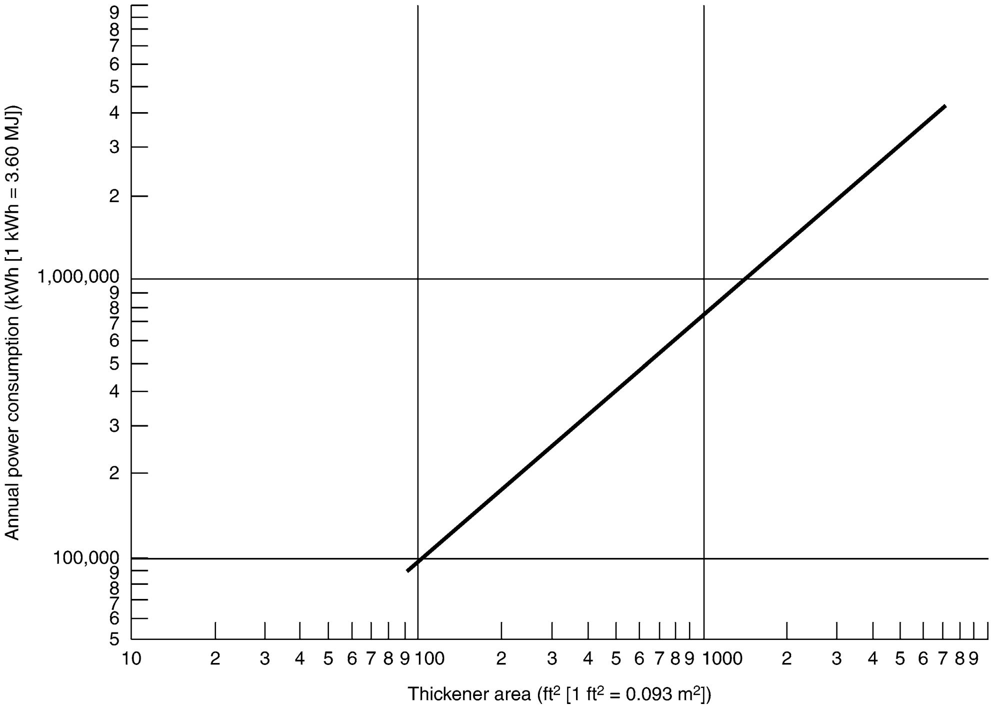

Fig. 8. Annual power consumption for continuous operating DAF thickeners (*Source:* US EPA).

wastewater. The air released to the atmosphere may strip volatile organic material from the sludge. The volume of sludge requiring ultimate disposal may be reduced, although its composition will be altered if chemical flotation aids are used. The air compressors will require shielding to control the noise generated.

8. PROCESS DESIGN CRITERIA AND PROCEDURES

8.1. Design Criteria

The following are the ranges of general design criteria for the DAF thickening process:

a. Pressure = 30–70 psig.
b. Effluent recycle ratio = 30–150% of influent flow.
c. Air to solids ratio = 0.02 lb air/lb solids.
d. Solids loading = 5–55 lb/ft^2/d (depending on sludge type and whether flotation aids are used).
e. Polyelectrolyte addition (when used) = 5–10 lb/t of dry solids.
f. Solids capture = 70 to >99%.
g. Total solids of unthickened sludge = 0.3–2%.
h. Total solids of thickened solids = 3–12%.
i. Hydraulic loading = 0.4–2 gal/min/ft^2.

Specific design criteria based on the types of sludges to be thickened are presented in Table 1.

8.2. Input Data of DAF Thickener Design

Two important input data must be gathered:

Table 5
Additional Average DAF Thickener and Clarifier Design Parameters

	Typical value	
Parameter	Thickening	Clarification
Air pressure (psig)	40–70	40–70
Effluent recycle (%)	130–150	30–120
Detention time (h)	3	0.25–0.5
Air-to-solids ratio (Ib air/lb solids)	0.005–0.06	–
Solid loading (lb/ft^2/d)		
Activated sludge (mixed liquor)	5–15	–
Activated sludge (settled)	10–20	–
50% primary + 50% activated	20–40	–
Primary only	To 55	–
Hydraulic loading (gpm/ft^2)	0.2–4	1–4
Detention time (min) (pressurizing tank)	1–3	1–3

Source: US EPA.

a. Wastewater (or sludge) flow (MGD).
b. Total suspended solids (TSS) concentration in the influent feed (mg/L).

Both average TSS concentration and variation in TSS concentration (i.e., TSS concentration range, [mg/L]) should be known.

8.3. Design Parameters

The following design parameters can be obtained from laboratory or pilot plant studies:

a. Air-to-solids ratio (A/S), from Table 5.
b. Air pressure (P [psig]).
c. Detention time in flotation tank (DTFT [h]).
d. Solids mass loading (SL [lb/ft^2/d]).
e. Hydraulic loading (HL [gpm/ft^2]).
f. Detention time in pressure tank (DTPT [min]).
g. Float concentration (X_f [%]).

8.4. Design Procedure for DAF Thickener With No Recycle (Direct Pressurization)

a. Select air-to-solids ratio (from laboratory or pilot plant studies; Table 5).
b. Calculate the required pressure:

$$A/S = \frac{1.3\,a\,(F\,P-1)}{X} \tag{2}$$

$$A/S = \frac{1.3\,a\,f\,(P-1)}{X} \tag{2a}$$

where A/S is the air-to-solids ratio (lb air/lb solids); a is the air solubility at standard conditions (mL/L); P is the absolute pressure (atm) = (psig + 14.7)/14.7; X is the suspended solids concentration in the feed (mg/L); psig is the gage pressure (lb/in^2); 1.3 is the mg weight of 1 mL of air; f is the factor of gas dissolution at pressure P (where $P > 1$ atm), fraction (usually 0.167–1); and F is the fraction of gas dissolution at pressure P (where $P > 2$ atm), fraction (usually 0.5 to 1). Since the1970s, design engineers have been successfully using Eq. (2) for high-pressure systems, such as DAF, where P is more than 2 atm (11–16,18,33). Eq. (2), however, is invalid for a low-pressure system where P is less than 2 atm. Recently Selke and his colleagues (37) have suggested a new factor (f) and Eq. (2a) to cover any pressure value more than normal atmospheric pressure of 1 atm.

c. Select a mass loading rate (or hydraulic loading rate) and calculate required surface area.

$$\text{SA} = \frac{(Q)\,(X)\,(8.34)}{\text{SL}} \tag{3}$$

$$\text{SA} = \frac{(Q)\,(1{,}000{,}000)}{(\text{HL})\,(60)\,(24)} \tag{4}$$

where SA is the surface area (ft^2); Q is the feed flow (MGD); X is the suspended solids concentration in the feed (mg/L); SL is the solids mass loading ($\text{lb/ft}^2\text{/d}$); and HL is the hydraulic loading (gpm/ft^2).

d. Use the larger of the two areas calculated in step c.
e. Select detention time and calculate volume of flotation tank.

$$\text{VOLFT} = (Q)\,(1/7.48)\,(1/24)\,(\text{DTFT})\,(1{,}000{,}000) \tag{5}$$

where VOLFT is the volume of flotation tank (ft^3); Q is the total flow (MGD); and DTFT is the detention time in flotation tank (h).

f. Calculate volume of sludge produced.

$$\text{VSP} = \frac{(Q)\,(X)\,\text{removal}\,(\%)}{(X_f)\,(\text{specific gravity})} \tag{6}$$

where VSP is the volume of sludge produced (gpd); Q is the total flow (MGD); X is the influent suspended solids concentration (mg/L); and X_f is the solids concentration in float (%).

8.5. Design Procedures for DAF Thickener With Recycle

a. Select air-to-solids ratio.
b. Assume pressure (40–60 psig).
c. Calculate P in atm = (psig + 14.7)/14.7.
d. Calculate recycle flow:

$$A/S = \frac{1.3\,a\,(F\,P-1)\,Q_r}{Q\,X} \tag{7}$$

$$A/S = \frac{1.3\,a\,f\,(P-1)\,Q_r}{(Q\,X)} \tag{7a}$$

where A/S is the air-to-solids ratio; a is the air solubility at standard conditions (mL/L); P is the absolute pressure (atmospheres); Q_r is the recycle flow (MGD); Q is the feed flow (MGD); and X is the influent suspended solids concentration (mg/L).

e. Select a SL rate (for thickening) or hydraulic loading rate (for clarification) and calculate surface area.

$$\mathrm{SA} = \frac{(Q)\,(X)\,(8.34)}{\mathrm{SL}} \tag{8}$$

$$\mathrm{SA} = \frac{(Q + Q_r)\,(1{,}000{,}000)}{(\mathrm{HL})\,(60)\,(24)} \tag{9}$$

where SA is the surface area (ft^2); Q is the feed flow (MGD); X is the influent suspended solids concentration (mg/L); SL is the solids mass loading rate (lb/ft^2/d); Q_r is the recycle flow (MGD); and HL is the hydraulic loading rate (gpm/ft^2).

f. Use the larger of the two areas calculated in step e.
g. Select detention time in the flotation tank and calculate the volume.

$$\mathrm{VOLFT} = (Q + Q_r)\,(1/7.48)\,(1/24)\,(\mathrm{DTFT})\,(1{,}000{,}000) \tag{10}$$

where VOLFT is the volume of flotation tank (ft^3); Q is the total flow (MGD); Q_r is the recycle flow (MGD); and DTFT is the detention time in flotation tank (h).

h. Select pressure tank detention time and calculate the volume of the pressure tank.

$$\mathrm{VOLPT} = (Q_r)\,(1/7.48)\,(1/24)\,(1/60)\,(\mathrm{DTPT})\,(1{,}000{,}000) \tag{11}$$

where VOLPT is the volume of the pressure tank (ft^3); Q_r is the recycle flow (MGD); and DTPT is the detention time in pressure tank (min).

i. Calculate volume of sludge.

$$\mathrm{VS} = \frac{(Q)\,(X)\ \text{removal}\ (\%)}{(X_f)\,(\text{specific gravity})} \tag{12}$$

where VS is the volume of sludge (gpd); Q is the feed flow (MGD); X is the influent suspended solids concentration (mg/L); and X_f is the solids concentration in the float (%).

8.6. Output Data of DAF Thickener Design

The output data of a DAF thickener design include:

a. Suspended solids concentration (mg/L).
b. Air-to-solids ratio.
c. Air pressure (psig).
d. Solids loading (lb/ft^2/d).
e. Hydraulic loading (gpm/ft^2).
f. Recycle flow (MGD).
g. Surface area (ft^2).
h. Volume of pressure tank (ft^3).
i. Volume of flotation tank (ft^3).
j. Pressure tank detention time (min).
k. Flotation tank detention time (h).

9. DESIGN AND APPLICATION EXAMPLES

9.1. Example 1. Design of a DAF Thickener With No Recycle

1. Select air-to-solid ratio (from laboratory or pilot plant studies).

From Table 5 ($A/S = 0.03$).

2. Calculate the required pressure
   a. Use Eq. (2) and assume $F = 0.5$

$$A/S = \frac{1.3\ a\ (FP-1)}{X} \tag{2}$$

where A/S is the air-to-solid ratio; lb air/lb solids = 0.03; a is the air solubility at standard conditions = 18.7 mL/L; F is the 0.5 (F ranges between 0.5 and 1 when P is more than 2 atm); and X is the suspended solids concentration in the feed = 200 mg/L.

$$0.03 = \frac{1.3\ (18.7)\ (0.5\ P-1)}{200}$$

Hence P is 2.5 atm.

   b. Use Eq. (2a) and assume $f = 0.167$

$$A/S = \frac{1.3\ a\ f\ (P-1)}{X} \tag{2a}$$

where A/S is the air-to-solid ratio (lb air/lb solids) = 0.03; a is the air solubility at standard conditions = 18.7 mL/L; f is the 0.167 (f ranges between 0.167 and 1 when P is >1 atm); and X is the suspended solids concentration in the feed = 200 mg/L

$$0.03 = \frac{1.3\ (18.7)\ (0.167)\ (P-1)}{200}$$

$$P = 2.5$$

where P is the absolute pressure (atmosphere) = (psig + 14.7)/14.7 = 2.5 atm; and psig is the gage pressure = 14.7 P – 14.7 = 14.7 (2.5) – 14.7 = 22 lb/in^2.

3. Select a mass loading rate (or hydraulic loading rate) from Table 5 and calculate surface area

$$\mathrm{SA} = \frac{(Q)\ (X)\ (8.34)}{\mathrm{SL}} \tag{3}$$

where SA is the surface area (ft^2); Q is the feed flow (1 MGD); X is the suspended solids concentration in the feed (200 mg/L); SL is the solids mass loading (10 lb/ft^2/d); and HL is the hydraulic loading (2 gpm/ft^2).

$$\mathrm{SA} = \frac{(1)\ (200)\ (8.34)}{10}$$

$$= 166.8\ \mathrm{ft}^2 \text{ based on solids loading}$$

$$\mathrm{SA} = \frac{(Q)\ (1{,}000{,}000)}{(\mathrm{HL})\ (60)\ (24)} \tag{4}$$

$$= \frac{(1)\ (1{,}000{,}000)}{(2)\ (60)\ (24)}$$

$$= 347.2\ \mathrm{ft}^2 \text{ based on hydraulic loading.}$$

4. Use the larger of the two areas calculated in step 3

Surface area = 347.2 ft^2 (say 348 ft^2).

5. Select the detention time and calculate the volume of flotation tank

$$\mathrm{VOLFT} = (Q)\,(1/7.48)\,(1/24)\,(\mathrm{DTFT})\,(1{,}000{,}000) \tag{5}$$

where VOLFT is the volume of flotation tank (ft^3); Q is the total flow (1 MGD); DTFT is the detention time in flotation tank (3 h); 7.48 gal = 1 ft^3; and 24 h = 1 d.

$$\mathrm{VOLFT} = (1)\,(1/7.48)\,(1/24)\,(3)\,(1{,}000{,}000)$$
$$= 16{,}711\ \mathrm{ft}^3.$$

6. Calculate the volume of sludge produced

$$\mathrm{VSP} = \frac{(Q)\,(X)\ \text{removal}\ (\%)}{(X_f)\ (\text{specific gravity})} \tag{6}$$

where VSP is the volume of sludge produced (gpd); Q is the total flow (1 MGD); X is the suspended solid concentration (200 mg/L); X_f is the solid concentration in float (5%); removal = 80%; and specific gravity = 1.05

$$\mathrm{VSP} = \frac{(1)\,(200)\,(80)}{(5)\,(1.05)}$$
$$= 3048\ \text{gpd}.$$

9.2. Example 2. Design of a DAF Thickener With Recycle

1. Select air-to-solids ratio (use 0.05).
2. Assume pressure = 40–60 psig (use 50 psig).
3. Calculate P in atmosphere = (psig + 14.7)/14.7 = P = (50 + 14.7)/14.7 = 4.4 atm.
4. Calculate recycle flow.
   a. Use Eq. (7) and assume $F = 0.5$

$$A/S = \frac{1.3\,a\,(F\,P-1)\,Q_r}{Q\,X} \tag{7}$$

where A/S is the air-to-solids ratio = 0.05; a is the air solubility at standard conditions (mL/L) = 18.7 mL/L; P is the absolute pressure (atmosphere) = 4.4 atm; Q is the feed flow (1 MGD); X is the influent suspended solids concentration = 200 mg/L

$$0.05 = \frac{1.3(18.7)(0.5\times 4.4-1)Q_r}{1\times 200}$$

Hence Q_r recycle flow = 0.343 MGD.

   b. Use Eq. (7a) and assume $f = 0.375$.

$$A/S = \frac{1.3\,a\,f\,(P-1)\,Q_r}{Q\,X} \tag{7a}$$

where A/S is the air-to-solids ratio = 0.05; a is the air solubility at standard conditions (18.7 mL/L); P is the absolute pressure (4.4 atm); Q_r is the recycle flow (MGD); Q is the feed flow (1 MGD); and X is the influent suspended solids concentration (200 mg/L)

$$0.05 = \frac{1.3\,(18.7)\,(0.375)\,(4.4-1)\,Q_r}{1\times 200}$$
$$Q_r = 0.343\ \text{MGD}.$$

5. Select a solids mass loading rate (for thickening) or hydraulic loading rate (for clarification) and calculate the surface area. Now only the hydraulic loading rate is selected for calculation of the surface area for the purpose of illustration.

$$\mathrm{SA} = \frac{(Q)\,(X)\,(8.34)}{\mathrm{SL}} \tag{8}$$

where Q is the feed flow (1 MGD); X is the influent suspended solids concentration (200 mg/L); and SL is the solids mass loading rate (10 lb/ft^2/d).

$$\mathrm{SA} = \frac{(1)\,(200)\,(8.34)}{10}$$

$$\mathrm{SA} = 166.8\ \mathrm{ft}^2 \text{ based on the solids loading}$$

$$\mathrm{SA} = \frac{(Q + Q_r)\,(1{,}000{,}000)}{(\mathrm{HL})\,(60)\,(24)} \tag{9}$$

where SA is the surface area (ft^2); Q is the feed flow (1 MGD); Q_r is the recycle flow (0.342 MGD); HL is the hydraulic loading rate (1.5 gpm/ft^2); 24 h = 1 d; and 60 min = 1 h.

$$\mathrm{SA} = \frac{(1 + 0.342)\,(1{,}000{,}000)}{(1.5)\,(60)\,(24)}$$

$$= 621.3\ \mathrm{ft}^2 \text{ (say 622 ft}^2\text{ based on hydraulic loading).}$$

6. Use the larger of the two areas calculated in step 5
Select a surface area = 622 ft^2.

7. Select detention time in the flotation tank and calculate the volume

$$\mathrm{VOLFT} = (Q + Q_r)\,(1/7.48)\,(1/24)\,(\mathrm{DTFT})\,(1{,}000{,}000) \tag{10}$$

where VOLFT is the volume of flotation tank (ft^3); Q is the influent flow (1 MGD); Q_r is the recycle flow (0.342 MGD); DTFT is the detention time in flotation tank (0.5 h); 7.48 gal = 1 ft^3, and 24 = 1 d.

$$\mathrm{VOLFT} = (1 + 0.342)\ (1/7.48)\ (1/24)\ (0.5)\ (1{,}000{,}000)$$
$$= 3738\ \mathrm{ft}^3.$$

8. Select pressure tank detention time and calculate volume of pressure tank

$$\mathrm{VOLPT} = (Q_r)\,(1/7.48)\,(1/24)\,(1/60)\,(\mathrm{DTPT})\,(1{,}000{,}000) \tag{11}$$

where VOLPT is the volume of pressure tank (ft^3); Q_r is the recycle flow (0.342 MGD); DTPT detention time in pressure tank = 3 min; 7.48 gal = 1 ft^3; 24 h = 1 d; and 60 min = 1 h.

$$\mathrm{VOLPT} = (0.342)\ (1/7.48)\ (1/24)\ (1/60)\ (3)\ (1{,}000{,}000)$$
$$= 95\ \mathrm{ft}^3.$$

9. Calculate volume of sludge

$$\mathrm{VS} = \frac{(Q)\,(X)\ \text{removal}\ (\%)}{(X_f)\ (\text{specific gravity})} \tag{12}$$

where VS is the volume of sludge (gpd); Q is the feed flow (1 MGD); X is the influent suspended solids concentration (200 mg/L); X_f is the solids concentration in float (5%); removal (%) is about 80; and specific gravity = 1.05

$$= \frac{(1)(200)(80)}{(5)(1.05)}$$

$$= 3048 \text{ gpd.}$$

9.3. Example 3. Complete DAF Thickening System Design

A designer has calculated that it will be necessary to thicken a maximum of 2700 lb (1225 kg)/d of waste sludge at 0.5–0.8% solids from a contact stabilization plant employing no primary clarification. The facility will have a sludge handling system consisting of a DAF thickener for the WAS, mechanical dewatering by belt press and composting. The treatment plant will be manned 8 h/d, 7 d/wk but dewatering operations will only take place 6 h/d, 5 d/wk. Thickening operation would take place 7.5 h/d, 5 d/wk. Waste sludge flow from the final clarifier would be continuous during the thickening operation that is, 7.5 h/d, 5 d/wk.

The designer has decided to provide polymer feed equipment for the DAF thickener to be used in emergency situations only. Polymers are not used in normal operation.

The designer has also decided to use a packed pressurization tank, which requires a relatively clean source of pressurized flow. Secondary effluent will be utilized.

9.3.1. Effective Surface Area

The maximum daily waste sludge production expected was given as 2700 lb (1225 kg) of WAS with a solids concentration of 5000–8000 mg/L.

The maximum net hourly load (actual amount of solids that must be captured and removed per hour by the thickener) is:

$$\frac{(2700 \text{ lb/d})(7 \text{ d/wk})}{(7.5 \text{ h/d})(5 \text{ d/wk operation})} = 504 \text{ lb/h } (228.6 \text{ kg/h})$$

The sludge being thickened is considered to be equivalent to a straight WAS even though primary solids are mixed with it. From Table 2, a value of 0.42 lb/ft^2/h (2.1 kg/m^2/h) is selected.

$$\frac{504 \text{ lb/h max net load}}{0.42 \text{ lb/ft}^2\text{/h loading rate}} = 1200 \text{ ft}^2 \ (108 \text{ m}^2)$$

Based on the solids loading rate (hydraulic loading rate needs to be checked), the maximum effective surface area required is 1200 ft^2 (108 m^2).

9.3.2. Feed Flow Rate Determination

Feed, pressurized recycle, thickened sludge, and subnatant must be calculated to determine pump size and piping requirements.

Both the gross solids load and minimum solids concentration must be known to calculate the feed flow rate. The gross solids load is the amount of solids that must be fed to the thickener in order for the system to capture and thicken the required net solids load. The maximum net hourly load has already been calculated to be 504 lb/h (228.6 kg/h). Since polymers are not to be used during normal operation, a capture efficiency of 85% is used (standard for the industry). The maximum gross solids load is then calculated as follows:

$$\frac{504 \text{ lb/h max net load}}{0.85 \text{ efficiency factor}} = 593 \text{ lb/h } (269 \text{ kg/h})$$

The minimum solids concentration expected is 5000 mg/L. The maximum feed flow rate can now be calculated as follows (note: 5000 mg/L = 0.5%):

$$\frac{593 \text{ lb/h}}{(0.005)\,(8.34)\,(60 \text{ min/h})} = 237 \text{ gpm } (897 \text{ L/min})$$

9.3.3. Pressurized Recycle Flow Rate

The design of the pressurized flow should be based on the maximum gross solids load expected from the DAF thickener. For this example, the maximum hourly gross solids load used was 593 lb/h (269 kg/h). After discussing the operating conditions with several DAF thickener equipment suppliers, the engineer designed for a maximum of 237 gpm (14.95 L/s), assuming 1 gal of liquid solids weighs 8.34 lb.

9.3.4. Thickened Sludge Flow Rate

The maximum hourly net solids load was 504 lb/h (228.6 kg/h). At the minimum 4% solids concentration, the expected flow rate can be calculated as follows:

$$\frac{5.04 \text{ lb/h}}{(0.04)\,(8.34)\,(60 \text{ min/h})} = 25.2 \text{ gpm } (1.59 \text{ L/s})$$

9.3.5. Subnatant Flow Rate

This rate is equal to the maximum total flow into the tank: 237 gpm (14.95 L/s) feed plus 237 gpm (14.95 L/s) recycle.

9.3.6. Hydraulic Surface Loading Rate

Based on solids loading, the minimum thickener surface area was calculated to be 1200 ft^2 (108 m^2). The total maximum flow rate (influent plus recycle) was calculated to be 474 gpm (1794 L/min). The maximum hydraulic surface loading rate would be:

$$\frac{474 \text{ gpm}}{1200 \text{ ft}^2} = 0.4 \text{ gpm/ft}^2 \ (0.27 \text{ L/s/m}^2)$$

The 0.4 gpm/ft^2 (16.3 L/min/m^2) is on the low side for a system that does not employ polymer addition. Under maximum conditions, polymer usage would be required.

9.3.7. Number of Units and Manufacturer's Recommendations

Only one unit will be used, with an adequate spare parts inventory to minimize down time. Several reputable manufacturers of DAF thickeners were contacted for their comments on the designer's calculations and proposed application.

NOMENCLATURE

a	Air solubility at standard conditions (mL/L)
A/S	Air-to-solids ratio (lb air/lb solids [kg air/kg solids])
C	Capital cost of DAF Thickener process in 2006 (USD)
DTPT	Detention time in pressure tank (min)
DTFT	Detention time in flotation tank (h)
HL	Hydraulic loading (gpm/ft^2 [L/min/m^2])
psig	Gage pressure (lb/in^2)
P	Absolute pressure (in atm)
Q	Flow MGD (ML/d)
Q_r	Recycle flow MGD (ML/d)
SA	Surface area ft^2 (m^2)
SL	Solids mass loading (lb/ft^2/d [kg/m^2/d])
t	Ton (English unit)
T	Tonne (Metric unit)
TSS	Total suspended solids concentration (mg/L)
VOLFT	Volume of flotation tank ft^3 (m^3)
VOLPT	Volume of pressure tank ft^3 (m^3)
VS	Volume of sludge per day gpd (L/d)
VSP	Volume of sludge produced per day (gpd [L/d])
X	Influent suspended solids concentration (mg/L)
X_f	Solids concentration in float (%)

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APPENDIX

United States Yearly Average Cost Index for Utilities[a]

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

[a] Extracted from US ACE (Tables Revised 31 March 2003) *Civil Works Construction Cost Index System Manual*, No 1110-2-1304, US Army Corps of Engineers, Washington, DC (PDF file is available on the Internet at http://www.nww.usace.army.mil/cost).

4

Centrifugation Clarification and Thickening

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CONTENTS

1. INTRODUCTION

Since 1920, centrifuges have been employed in the US wastewater treatment industry. However, the early centrifuges did not perform well because of their complex design and poor mechanical reliability. In the early 1960s, a number of innovative designs were developed, which solved most of the operational problems. Sludge dewatering or thickening, and disposal of the residual solids, are significant costs of the waste treatment operation. These steps are especially important if the final disposal option is incineration. The incinerator will be more efficient if more water is removed prior to incineration. If the sludge has to be transported to a landfill, volume reduction owing to dewatering will significantly improve the overall economics (1–49). The sludge concentration methods of clarification, filtration, settling, flocculation, thickening, and centrifugation were evaluated. The choice of a dewatering technique is based on the feedstock, device design and function, and engineering specifications dependent on the intended outlook of the final products (47–48). Filter belt presses, solid-bowl centrifuges, and membrane filter presses were analyzed for their ability to dewater sludge according to their capacity, polymer consumption of dewatered dry solids, suspend solids in the reject water, and economics (49).

The idea of "bound water" regarding sludge dewatering was reclassified (50). Sludge-polymer conditioning and mechanical dewatering that only eliminates the bulk

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and interstitial waters and not the vicinal water was proposed as the cause behind the occasional ineffective dewatering. In Taiwan, a technique using both vacuum filtration and centrifugal settling tests, together for bound water measurements was suggested (51). In 1997, Wench and Otte-Witte (52) reviewed the high-tech methods in biosolids and sludge dewatering with centrifuges together with the design and process of centrifugation, process engineering, optimization, dewatering effectiveness, and monetary impacts. Thermally conditioned wastewater biosolids have the ability to become effectively dewatered using a centrifuge (53). Additional data on the principles, applications, performance, and costs of centrifuges can be found in the literature (54–70).

Although, there are a number of dewatering methods employed in the wastewater treatment industry, this chapter focuses on centrifugal separation and sedimentation centrifuges in particular. Three types of sedimentation centrifuges will be discussed, namely, the solid-bowl, basket, and disc centrifuges (1–4). Included are advantages and disadvantages, a general description, and principal features of these centrifuges. The performance of the centrifuges with respect to municipal wastewater treatment sludges, coal-washing slimes, electroplating waste treatment sludges, paper and pulp mill sludges, and biological sludges is also addressed. To conclude, a guide to selection of centrifuges is also given in this chapter.

2. PRINCIPLES

The settling rate of solids in a liquid can be greatly increased by replacing gravitational force by a strong centrifugal force (1). This separation is based on the density difference between solids and liquids (or between two liquids). The lighter liquid remains near the center of rotation and overflows over weir. The heavier solids move to the periphery of the bowl and are discharged either continuously or intermittently. In a perforated basket centrifuge, the liquid percolates through a solid cake at the periphery. Stoke’s Law for the settling rate is described in another chapter entitled “Sedimentation.” The sedimentation equation can be used as a general guideline, and more exact formulae can be developed through laboratory experiments and scale-up procedures.

Scale-up with centrifuges may be difficult. Although the geometric configuration of a centrifuge may be similar in model and full scale, the actual flow regime may be different. Trawinski (4) gives an overview of solid–liquid separation processes and can be referred to for detailed information. Bates College’s Department of Biology (66) presents more facts on centrifugation basics. The University of Texas A&M University, Natural Toxins Research Institute (67) presents detailed theory and principles on centrifugation.

3. TYPES OF WATER ASSOCIATED WITH SOLID PARTICLES

During the sludge dewatering process, it is important to identify different types of water (5–7) associated with a sludge particle such as:

a. Bulk or free water.
b. Micropore water.
c. Colloidally bound water.
d. Chemisorbed water.

Success of the dewatering depends upon the percentage of each of these waters associated with solid sludge particles. For example, the waters may be present in sludge as follows:

a. Bulk water 40%.
b. Micropore water 40%.
c. Colloidally bound water 20%.
d. Chemisorbed water 10%.

Conventional solid–liquid separation techniques such as centrifugation can separate only the bulk water present and not the other types. Hence, the solids concentration achievable by these devices is rather modest. Chemical conditioning agents such as polymers are used to enhance the solids concentration.

4. TYPES OF CENTRIFUGES

Various types of centrifuges are employed in the wastewater treatment industry. However, in this discussion, only three centrifuges will be discussed. They are as follows:

a. Basket centrifuge.
b. Solid-bowl centrifuge.
c. Disc centrifuge.

4.1. Basket Centrifuge

This type of centrifuge has been employed in the treatment of both municipal wastewater and industrial sludges. The earliest device was developed by Herman Schafer in 1902 to dewater primary sludges in Cologne, Germany (7). The basket centrifuge is especially suited for smaller plants in the range of 2–4 million gal/d. A basket centrifuge is basically a rotating vertical chamber that has a weir at the top. It is a semicontinuous device and sludge is fed at the bottom. A schematic diagram of the unit is shown in Fig. 1. Solids move to the chamber wall and the clarified liquid flows over the weir and is taken out. Centrifuges operate in G units, which is the "ratio of centrifugal acceleration to the acceleration of gravity." The centrifugal field applied in this unit varies anywhere from 6000 to 10,000 G on bench-scale or up to 1500 G on commercial scale. Consolidated cake is removed intermittently. The basket decelerates and accelerates depending upon whether the solids are removed or the slurry is fed. Some of the operating characteristics of a basket centrifuge are as follows:

a. Typical cycle time 6–8 min.
b. Solids achievable 10–25% wt.
c. Feed rate 50–75 gal/min.

Lui and Liptak (42) state that the basket centrifuge is most appropriate for soft or small solids that are problematic to filter and waste that fluctuates in concentration and solids characteristics. The design has a three-point casing suspension and a hydraulic or electric motor powers that provides water supply to it. A 12 or 14 in. diameter bowl is typically used for testing; however, the diameter for commercial bowls is 30–48 in. Sludge feed enters at the top, travels near the bottom of the bowl and flows axially. The clarified effluent is released at the top. Suspended solids settle on the bowl wall and ultimately block

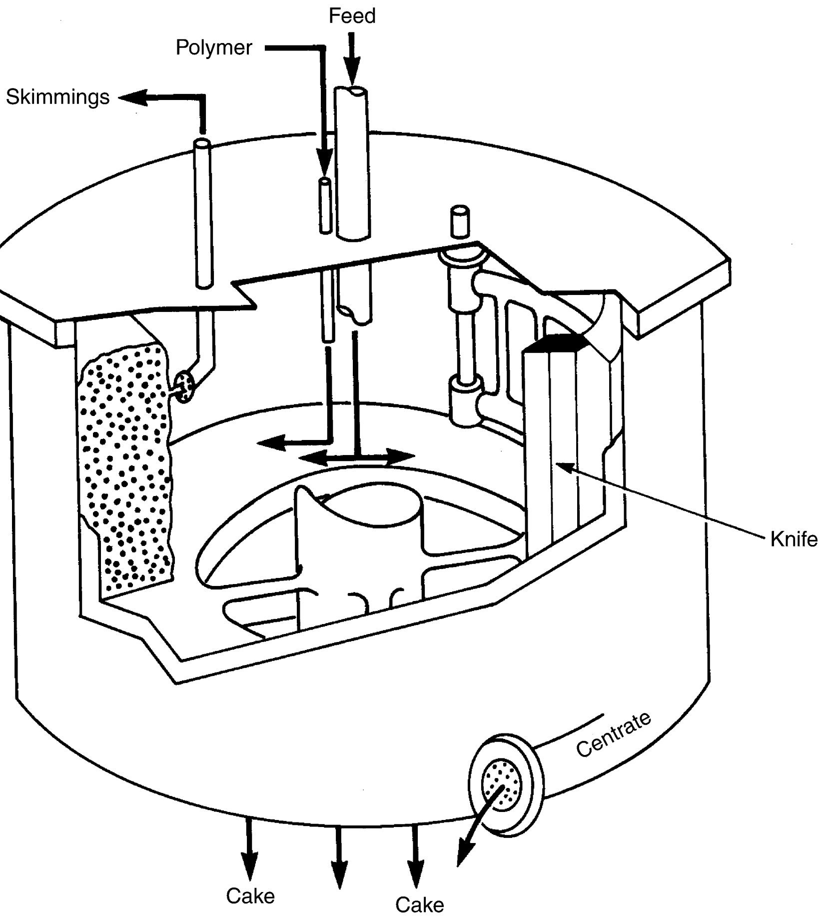

Fig. 1. Schematic of a basket centrifuge (*Source:* US EPA).

clarification. Maintenance is minimal owing to the fact that the basket centrifuges operates at a lower speed than other centrifuges. The bearing life is 1–3 yr. Lubrication is easy, little power is consumed, and the parts seldom need replacement. The advantages and disadvantages of basket centrifuges are given in Table 1 in relation to solid-bowl and disc centrifuges.

4.2. *Solid-Bowl Centrifuge*

Sometimes the solid-bowl centrifuge is referred to as a scroll or decanter centrifuge. This unit can be operated continuously and handles up to 100 t/h of cake product quantities of material. A schematic diagram of the unit is shown in Fig. 2. A screw-type conveyor for transporting the solids rotates at a slower or faster rate than the bowl itself. In a solid-bowl centrifuge, there are two different modes of contacting: countercurrent and concurrent. Sludge is fed through a feed pipe and the solids move to the bowl wall. Separated solids are slowly moved out through the back-drive. Centrate is drawn off and the cake proceeds for removal. The separated solids are not disturbed by the feed slurry (7–10). Advantages and disadvantages of solid-bowl centrifuge are given in Table 2.

Table 1
Advantages and Disadvantages of a Basket Centrifuge

Advantages	Disadvantages
Can handle small throughputs	Semicontinuous
Sludge difficult to dewater can be handled	Low solids concentration in product
Grit in the feed does not pose problem	Poor solids capture
Good flexibility	Low capacity
	G limitations

Source: From refs. 8 and 9.

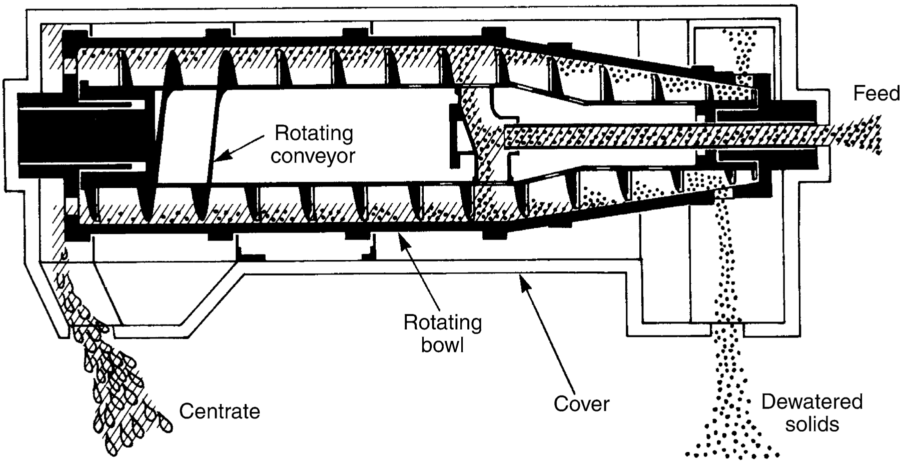

Fig. 2. Schematic of typical solid bowl decanter centrifuge (*Source:* US EPA).

Table 2
Advantages and Disadvantages of a Solid-Bowl Centrifuge

Advantages	Disadvantages
Little operator attention	Very noisy
Ability to handle variety of sludges	Poor supernatant quality
Larger throughputs compared to basket	High wear
Lower polymer dosages required	High maintenance costs
Easy startup and shutdown	Greater sensitivity to feed fluctuation
Can separate out fines	
Can handle dilute sludges	

Centrifuges are operated at a range of speed depending on the size and purpose to be achieved by the process. Thus, centrifugal machines can be operated at either high speed or low speed (10). A typical plot of variation of cake solids concentration with centrifugal acceleration for different types of wastewater sludges is shown in Fig. 3. Higher speeds permit smaller bowl diameters and effects greater solids capture and drier cakes. However, high-speed machines consume more power and generate more noise. Also, abrasion at high speed levels increases maintenance and replacement costs.

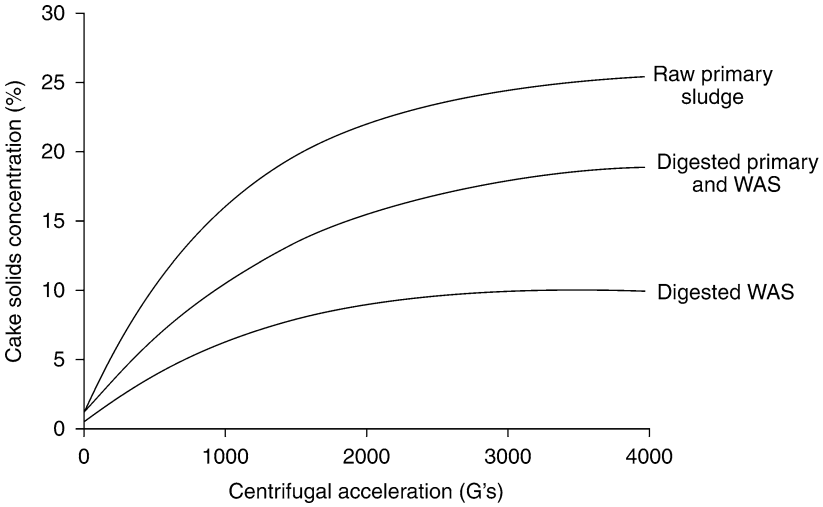

Fig. 3. Typical plot variation of cake solids concentration with centrifugal acceleration (*Source:* US EPA).

Machines operated at low speed levels consume less power and generate less noise but may result in lower solids capture and wetter cakes. One approach is to introduce the feed sludge as far as possible from the cake discharge end in order to permit longer solids residence times, thus enabling lower speed levels. In another approach, the centrifuge is redesigned to optimize the use of polymer and is operated at moderately high speed levels. Figure 4 shows the relative influence of one process variable as a function of feed solids content whereas imperforate basket centrifuge holding all other process variables constant.

The high-speed decanter centrifuge (11–13) is equipped to resist abrasion from coal particles by attaching tungsten carbide tiles (some manufacturers use ceramic tiles), thus enabling it to operate efficiently at high speed levels. At feed rate of 120–180 gal/min, feed concentration of 30–50% with 80% smaller than 325 mesh, the Super-D-Canter Model P-5400, which is a typical high-speed decanter centrifuge, achieves more than 99% recovery with less than 1 lb/t of polymer for refuse containing 45–50%. The Water Environment Federation (WEF) Manual of Practice offers considerations for designing a cost effective solid-bowl centrifuge (43). One recommendation is that a polymer addition is usually needed for solid-bowl centrifuge dewatering. The design must permit sludge feed directly into the centrifuge and upstream of the centrifuge. The design should take account of space requirements of the feed system, testing of various polymer concentrations, and flexibility to allow future modifications.

Design considerations also include the requirements for cake discharge and its conveying system. Conveyors are able to carry large quantities of cake; however, they are normally a housekeeping task and need special space conditions. Pumps are typically used for conveying the cake, but may not have the ability to convey more recent high-cake solid concentrations.

Centrate normally travels to the head of a plant, and a recycling pump station often transports the centrate to a specified location. Therefore, centrate piping has to be sized

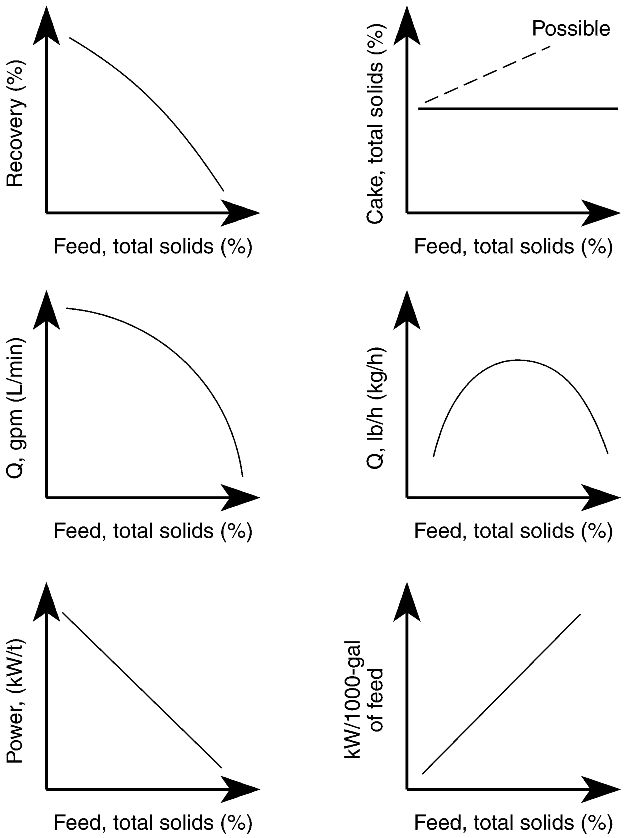

Fig. 4. Relative influence of one process variable as a function of feed solids content for imperforate basket centrifuge holding all other processes as variables constant (*Source:* US EPA).

and sloped appropriately to avoid centrate backups. Polymers frequently produce a foam or froth, as a result; a froth spray is usually needed. A sampling station can be utilized to detect any major centrate loads on the plant components. The area and space for a large machine is approx 40 m^2, which is significantly less space than the majority of mechanical-dewatering equipment with the same capacity. Most designs lack adequate space, operational flexibility, and unit redundancy.

4.3. Disc Centrifuge

In the early 1940's, disc centrifuges were first used in the United States. Presently, there are more than 100 such machines in a number of municipalities. A schematic view of the Disc no. 331e centrifuge is shown in Fig. 5. The sludge is fed from the top; however, feed from the bottom of the unit is also possible. The feed passes down to the center where a rotor distributes the feed sludge, filling the outer chamber. Advantages and disadvantages of the disc centrifuge are shown in Table 3.

Heavier solid particles settle toward the wall of the rotor because of the centrifugal field. The collected liquid and the lighter solids flow inside the disc stack. Generally the gap maintained between the discs is about 0.05 in., which means that the lighter

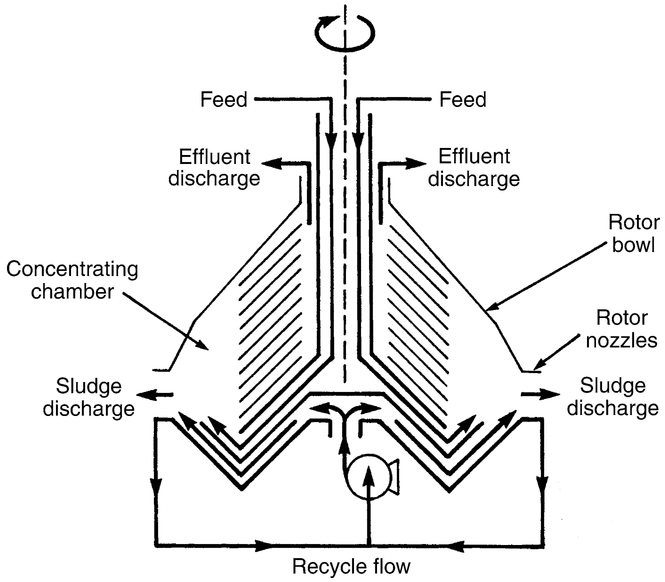

Fig. 5. Schematic of a disk nozzle centrifuge (*Source:* US EPA).

Table 3
Advantages and Disadvantages of a Disc Centrifuge

Advantages	Disadvantages
Can handle large throughputs	Large particles, fibers cause clogging
Good quality concentrate can be produced	High maintenance requirements
Applicable to sludges having finer particles	Skilled operators required

particles have to travel a short distance before they settle on the surface of the disc. The supernatant liquid passes through the stack into an overflow chamber and is further discharged into an effluent line (7).

The disc centrifuge is particularly suitable for the thickening of excess activated sludge or alum sludges that does not contain coarse solids. It can manage high feed rates with efficient clarification for extremely fine particles with low concentrations. The disc centrifuge must be used specifically for continuous extended running; the bowl must be cleared of solids before the centrifuge can be restarted after a shutdown. Little maintenance is needed because bearings in smaller units are grease-packed, while larger units have circulating oil or spray-mist systems. For abrasive applications, some areas require hard surfacing, however; nozzle bushings are replaceable (42).

Units can be online for as long as 8 wk when screening of feed is used before centrifugation. Inline self-cleaning strainers are usually acceptable. Because the feed can vary, control is suggested for consistency of thickened sludge. This control can be manual or fully automated, which uses a viscosity-sensing loop to indicate the sludge discharge consistency and moderate the recycling rate (42).

5. PERFORMANCE OF CENTRIFUGES IN SLUDGE DEWATERING

5.1. Separation of Municipal Wastewater Treatment Plant Sludges by Centrifugation

Several environmental problems are associated with the handling of the Municipal Wastewater Treatment Plant Sludges (MWTP sludges; 12). These sludges are odorous, potentially hazardous, and produce combustible gases in landfills. MWTP sludges typically contain less than 1% of weight solids. Hence dewatering/drying is required to render these sludges suitable for land filling or incineration. A variety of centrifuges are used to dewater MWTP sludges in the United States.

The performance of centrifuge is measured by the following parameters:

a. Final solids concentration.
b. Solids recovery in centrate.
c. Amount of polymer and chemicals used to treat a dry ton of sludge.

5.1.1. Basket Centrifuge Performance

Table 4 lists the typical performance data of a basket centrifuge. Results show that with raw primary sludge, 25–30% solids cake can be achieved; however, with raw waste-activated sludge, the performance drops drastically. For autogeneous combustion of these sludges, approx 30–35% solids is required. The centrate quality in the case of mixed anaerobically digested sludge and waste-activated sludge is very poor.

5.1.2. Solid-Bowl Centrifuge Performance

The dewatering performance of the solid-bowl centrifuge is good for a broad range of sludges (14). Table 5 shows the typical data for different types of MWTP sludges. A clear concentrate was produced at higher acceleration force. As the bowl speed is increased, the polymer dosage requirements decrease and the solids concentration increases. Generally, the cake solids concentration levels are higher in a solid bowl than in a basket centrifuge for equivalent amounts of polymer addition. Raw primary sludge can be dewatered up to 25–36% while with waste-activated sludge only 8–12% solids is achieved. Polymer addition amounts in the case of anaerobically digested sludges is rather significant.

5.1.3. Disc Centrifuge Performance

Disc centrifuges have been used extensively to dewater MWTP sludges in the United States. It has been reported that a wide variety of problems in the beginning were attributed to clogging of the lines. Generally, some kind of pretreatment is recommended to remove fibrous material and large particles from the feed stream. Typical performance of a disc centrifuge is given in Table 6. It should be noted that polymer was not added and the solids level typically increased from 1% solids to 5 or 6% solids. This improvement is generally considered good because 80% of the liquid is removed as a clear concentrate even without the addition of a flocculating agent. Research is needed to generate the energy requirement data for disc centrifuges. Table 7 has energy requirement data for other centrifuges, except disc centrifuges.

5.1.4. Costs and Energy Requirements of Solid-Bowl and Basket Centrifuges

Capital, operating, and maintenance costs are presented in Figs. 6–9 for a basket centrifuge and a high-G solid-bowl centrifuge. Capital or construction costs include

Table 4
MWTP Sludge Performance Data for a Basket Centrifuge

Sludge type	Feed solids concentration (%)	Average cake solids concentration (%)	Dry polymer required g/kg feed solids (lb/t)	Recovery based on centrate (%)
Raw primary	4–5	25–30	1–1.5 (2–3)	95–97
Raw trickling filter (rock or plastic media)	2–3	9–10	0	90–95
		10–12	0.8–1.5 (1.5–3)	95–97
Raw waste activated	0.5–1.5	8–10	0	85–90
		12–14	0.5–15 (1–3)	90–95
Raw primary plus rock trickling filter (30:70)	2–3	9–11	0	95–97
		7–9	0.8–1.5 (1.5–3)	94–97
Raw primary plus waste activated (50:50)	2–2	12–14	0.5–1.5 (1–3)	93–95
Raw primary plus rotating biological contactor (60:40)	2–3	20–24	0	85–90
		17–20	2–3 (4–6)	98+
Anaerobically digested primary plus waste-activated (50:50)	1–2	12–14	0	75–80
		10–12	0.8–1.5 (1.5–3)	85–90
		8–10	2–3 (4–6)	93–95
Anaerobically digested	1–3	8–11	0	80–95
		12–14	0.5–1.5 (1–3)	90–95

Source: From ref. 15.
Skimming losses, if any, have not been use in calculating recovery.

purchased equipment structural work, site preparation, piping, instrumentation, electrical installation, and the necessary labor. The costs are represented in 1981 USD. For basket centrifuges, the construction cost is somewhat flat up to 30,000 gpd and increases rapidly for higher capacity machines. The construction cost of a high-G solid-bowl centrifuge increases exponentially from 10 to 1000 gal/min. Annual operating and maintenance costs include costs such as labor, energy, chemicals, and so on. Electrical energy includes heating, cooling, and lighting. Process energy includes energy for process equipment such as motors and drives. Energy required to dewater a ton of sludge will have an important bearing on the choice of a dewatering device to be used in a particular project. Direct energy is the actual energy utilized by the device, while the indirect energy is associated with the production of polymers and other chemical reagents. Table 7 provides information on both these subjects.

5.2. Separation of Pulp and Paper Sludges by Centrifugation

In the last few decades, the pulp and paper industry in the United States has made significant progress toward reducing the biochemical oxygen demand and total suspended

Table 5
Typical MWTP Performance Data for a Solid-Bowl Centrifuge

Sludge type	Feed solids concentration (%)	Average cake solids concentration (%)[a]	Dry polymer required g/kg feed solids (lb/t)	Recovery based on a centrate (%)
Raw primary	5–8	25–36	0.5–2.5 (1.5)	95–97
	28–36	0	35–45 (70–90)	
Anaerobically digested primary	2–5	28–35	3–5 (6–10)	98+
	9–12	30–35	0	65–80
		25–30	0.5–1.5 (1–3)	82–92
Anaerobically digested primary irradiated at 400 krad	2–5	28–35	3–5 (6–10)	95+
Waste activated	0.5–3	8–12	5–8 (10–15)	85–90
Aerobically digested waste activated	1–3	8–10	1.5–3 (3–6)	90–95
Thermally conditioned:				
Primary + waste activated	9–14	35–40	0	75–85
	13–15	29–35	0.5–2 (1–4)	90–95
Primary + trickling filter	7–10	35–40	0	60–70
		30–35	1–2 (1–4)	98+
High lime	10–12	30–50	0	90–95
Raw primary + waste activated	4–5	18–25	1.5–3.5 (3–7)	90–95
Anaerobically digested (primary + waste activated)	2–4	15–18	3.5–5 (7–10)	90–95
	4–7	17–21	2–4(4–8)	90–95
Anaerobically digested (primary + waste activated + trickling filter)	1.5–2.5	18–23	1–2.5 (2–5)	85–90
		14–16	6–8 (12–15)	85–90
Combined sewer overflow treatment sludge	Highly variable			

Source: From ref. 15.

[a] Assumes skimming of cake.

MWTP stands for the municipal wastewater treatment plant.

solids in the wastewaters discharged to rivers. It has been estimated that about 2.15 billion gal of wastewater sludge containing 6.04 million lbs of biochemical oxygen demand are treated daily by US. Pulp and Paper Industry (16). This treatment yields large quantities of primary and secondary sludges. Approximately 2.1 million t/yr of primary sludge is produced (1). Pulp and paper mill sludge consists mainly of fiber, fillers, coating clays, and several other minor impurities. Table 8 shows the normal sludge composition from different types of paper and pulp mills. Ash content (50–70%) significantly affects the performance of centrifuges (17).

Major dewatering processes used in the paper and pulp industry are vacuum filters, centrifuges and belt presses. Table 8 indicates different types of dewatering equipment

Table 6
Typical Performance of a Disc Nozzle Centrifuge

References	Capacity (gal/min)	Feed solids (%)	Underflow solids (%)	Solids recovery (%)	Polymer pounds per dry ton of solids
5	150	0.75–1	5–5.5	90+	None
5	400	–	4	80	None
5	50–80	0.7	5–7	93–87	None
5	60–270	0.7	6.1	97–80	None
24	66	1.5	6.5–7.5	87–97	None
60	200	0.75	5	90	None

Source: From ref. 7.
1 gpm = 3.78 L/min.
1 lb/t = 0.5 kg/t, where 1 t = 2000 lb; 1 T = 1000 kg.

Table 7
General Ranges of Direct and Indirect Energy Requirement for Sludge Dewatering

Process	Direct electricity per dry solids (kWh/kg)	Direct electricity per dry solids (kWh/t)	Polymer dosage (g/kg)	Indirect electricity per dry solids (kWh/t)	Indirect energy per dry solids (kWh/t)
Basket centrifuge	0.105–0.140	(90–120)	3	0.0007	(0.6)
Solid bowl centrifuge					
Low speed	0.035–0.070	(30–60)	4	0.009	(0.8)
High speed	0.070–0.105	(60–90)			

Source: US EPA.
Digested 50:50 Mixture of Primary + WAS at 3% solid feed.

used in the pulp and paper industry. Table 9 clearly indicates that vacuum filters are most widely used by the industry (16,17). Table 10 shows dewatering methods used by mixed primary and secondary sludges. Addition of chemical conditioning agents plays a major role in the treatment of mixed sludges. Vacuum filters again seem to be used widely in this industry (19).

Table 11 shows typical data for activated sludges from paper mills for three different types of centrifuges (20). Disc centrifuges in series with solid-bowl centrifuges appear to provide higher solids concentration and solids recovery than the other types (20–22). Because polymer is the principal operating cost, the strategy is to minimize the polymer consumed while maximizing the cake solids and solids recovery.

5.3. Separation of Electroplating Wastes by Centrifugation

Metal finishing plants produce a wide variety of process wastewaters (23–27) containing heavy metals. These wastewaters are treated to reduce the solubility of the metals and for further separating the precipitated metal hydroxides. This treatment

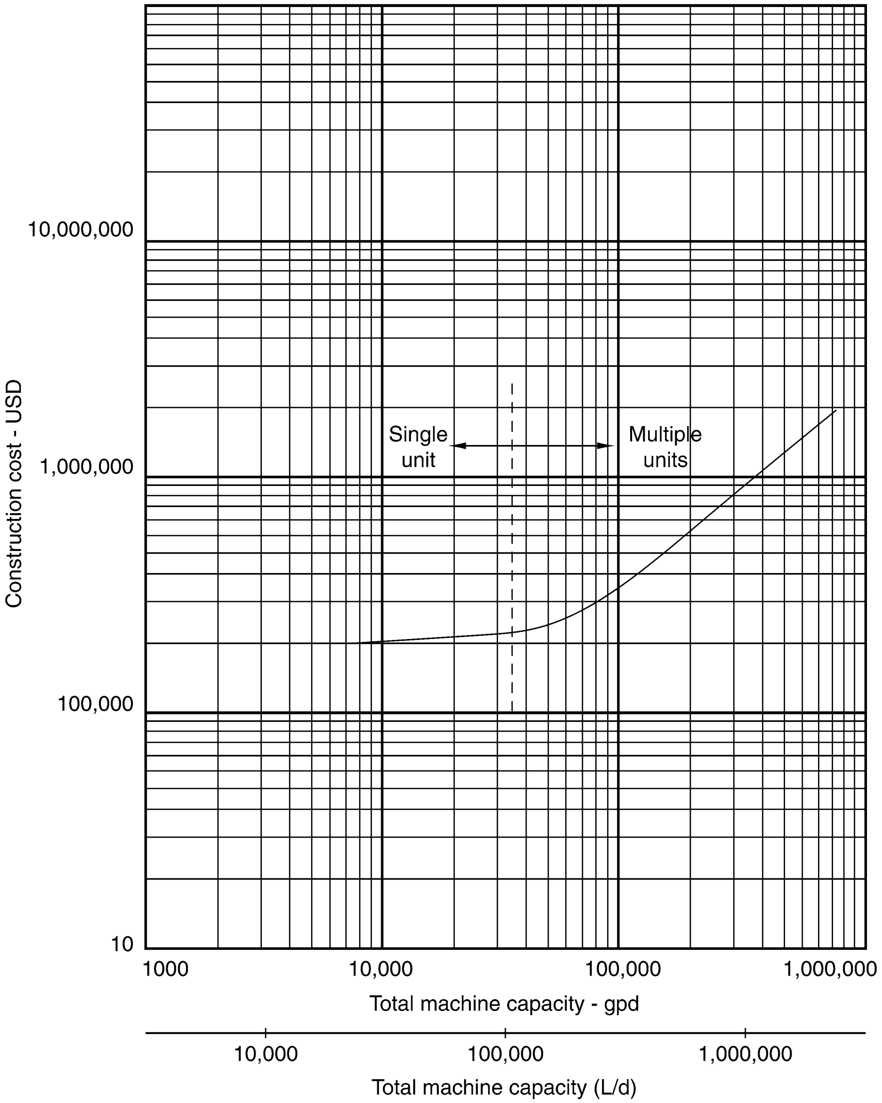

Fig. 6. Construction costs for basket centrifuges (*Source:* US EPA, 1982).

yields a solid waste containing high levels of toxic materials. Typical composition of an electroplating wastewater and the volumes at 3 and 25% solids by weight is given in Table 12. Typical performance of a basket centrifuge is given in Table 13. This sludge should be treated properly and disposed of in an environmentally acceptable manner.

The typical solids concentration of the sludge is 3% weight; hence, dewatering is an economical way of increasing the solids concentration of the sludge. This increment of solids concentration is done before it is sent to a landfill (28) or of the sludge treated in an incinerator. The strong centrifugal force greatly speeds up the settling process and consolidates the cake. This mechanism is important because the sludges that are obtained from the electroplating industry are compressible. Figure 10 shows the unit cost and horsepower (hp) rating for centrifuges.

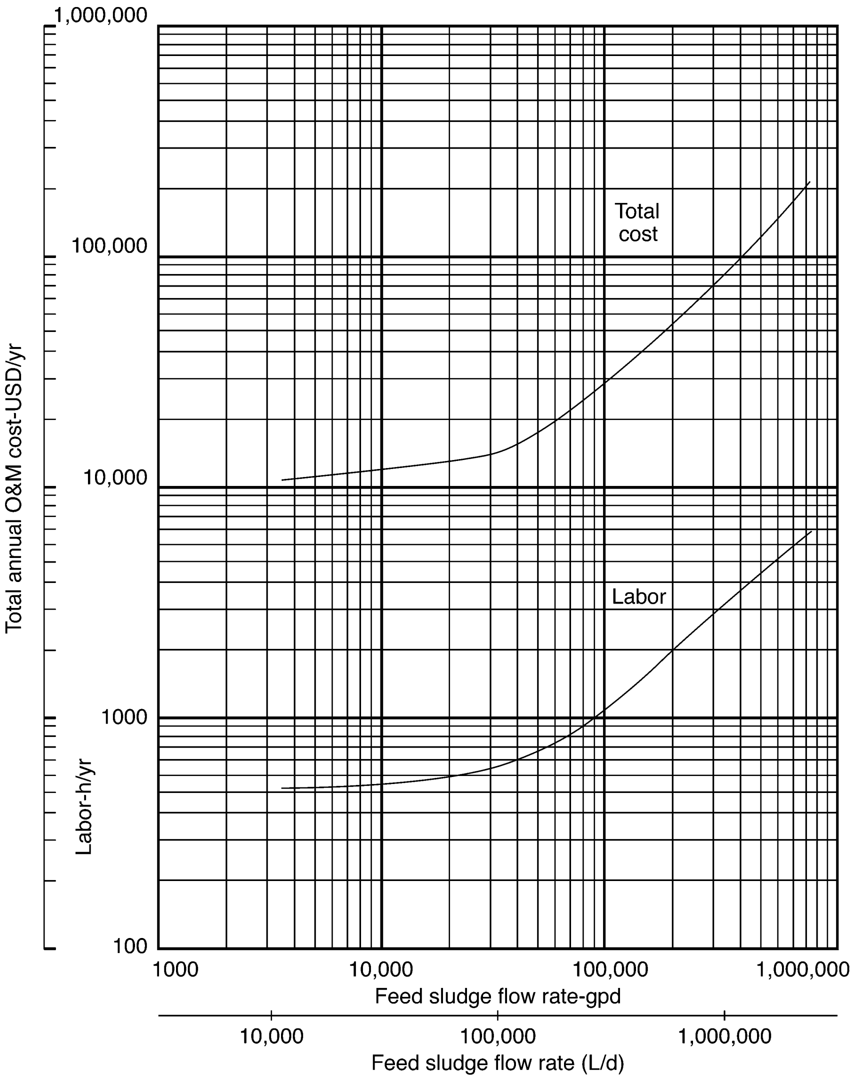

Fig. 7. Basket centrifuges—labor and total annual operation and maintenance costs (*Source:* US EPA, 1982).

A typical dewatering system with auxiliary equipment and its annual sludge disposal costs is shown in Fig. 11. The typical cost of installing and operating a basket centrifuge for an electroplating sludge is shown in Table 14. The installed cost for up to 150 gal/h feed is 48,000 USD, while for 200 gal/h feeds, the installed cost increases 3.5 times. The increase in the annual operating cost is marginal.

5.4. Separation of Coals and Refuse by Centrifugation

Mined coal is usually unsuitable for efficient utilization because of its high shale content. Beneficiation techniques use water for cleaning. Beneficiated coal must be dewatered to improve its heating value and reduce its bulk, while the refuse water must be

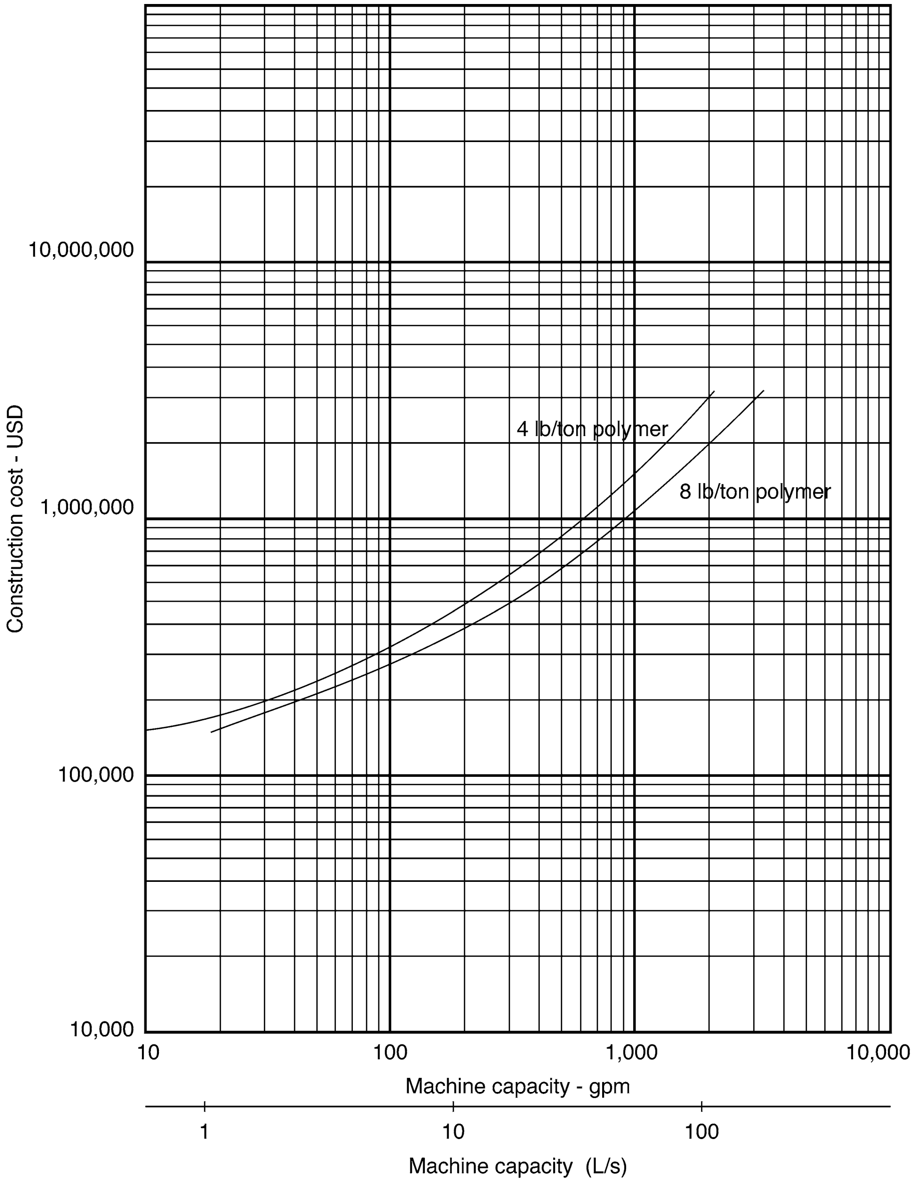

Fig. 8. Construction costs for high "G" solid-bowl centrifuges (*Source:* US EPA, 1982).

cleaned for recycling and recovery of fines. Coal and refuse coarser than 2.5 cm is dewatered by screens and centrifugal dewatering is used for coal and refuse between 590 μm and 2.5 cm (30). Screen-bowl and solid-bowl centrifuges with a scroll conveyor are commonly used for coal dewatering. Solid bowls can recover solids as small as 10 μm (29). A solid-bowl centrifuge manufacturer reports a solids product with 30–35% moisture and a clean liquid product, with feed slurry, which had 75–85% of solids less than 325 mesh size. The units had a capacity from 2.5 to 30 t. Manufacturers claim that the centrifuge requires half the hp of comparable vacuum disc filters. Typical centrifuge performance data is shown in Table 15.

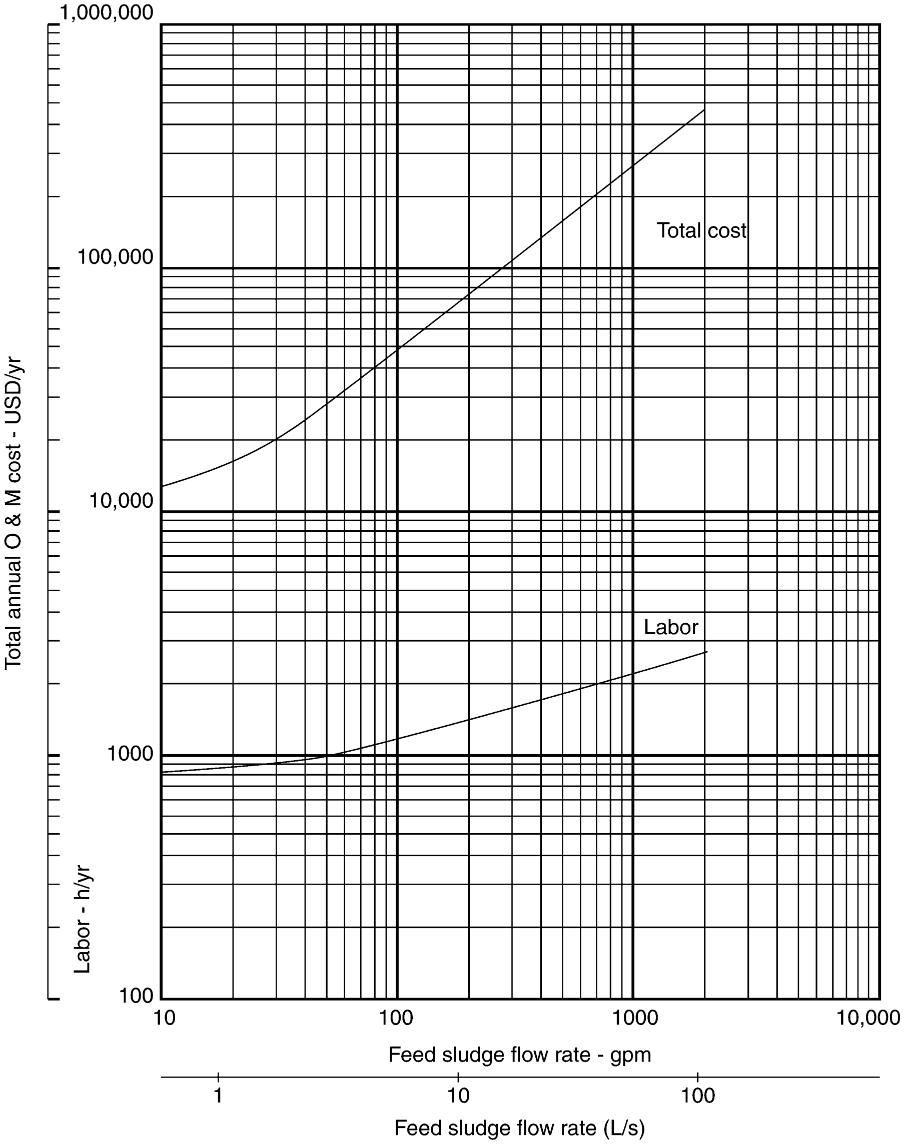

Fig. 9. High-G solid-bowl centrifuges—labor and total annual operation and maintenance costs (*Source:* US EPA, 1982).

Table 8
Normal Sludge Composition

Type	Solids (%)	Ash content (%)
Board mills	2–10	50–70
Chemical pulp	1–10	20–50
Deink pulp	3–10	25–60
Ground wood	2–5	1–20
Paper mills	1–5	50–70

Source: Design carried for ref. 18.

Table 9
Type of Dewatering Devices Used in the US Pulp and Paper Industry

Dewatering technology	Number of mills using technology	Total (%)
Vacuum filters	71	29
Lagoon	67	27.3
Belt press	35	14.3
Centrifuges	23	9.4
Screw press	16	6.5
V. presses	8	3.3
Flotation	5	2
Filter presses	4	1.6
Heat drying	1	0.4
Other	15	6.1
	245	99.9

Table 10
Number of Mills Following Various Approaches to Sludge Dewatering

	Report of mixed sludges mills using preconditioning step				
Dewatering technology	None	Gravity thickening	Chemical conditioning	Thickening plus chemicals	Total
None	–	1	0	0	1
Centrifuge	2	0	3	0	5
Vacuum filter	10	3	6	5	24
Belt press	3	0	11	7	21
Pressure filter	0	0	4	1	5
Mechnical press only	0	0	0	1	1
Basin drying	7	0	0	0	7
Basin disposal	2	0	0	0	2
Total	24	4	24	14	66

Source: From ref. 19.

Table 11
Centrifugal Dewatering of Activated Sludge From Paper Mills

Type of centrifuge	Polyelectrolytes (lb/t)	Solid recovery (%)	Cake solids (%)
Solid-bowl scroll	10–15	75–85	13–19
Disk	None	80–90	5–7
Solid-bowl basket	None	80–90	8–12
Disk in series with solid-bowl scroll	10–15	75–85	20–22

Source: From ref. 20.

Table 12
Sludge Generated and Sludge Disposal Volume for Electroplating

		Sludge (gal/lb metal precipated)	
Waste metal components	Dry solids (lb/lb metal precipitated)[a]	Generated (at 3% solids by weight)	Dewatered (to 25% solids by weight)
Aluminum	2.89	11.2	1.14
Cadmium	1.3	5	0.51
Chromium	1.98	7.7	0.79
Copper	1.53	5.9	0.6
Iron	1.61	6.2	0.63
Nickel	1.58	6.1	0.62
Zinc	1.52	5.9	0.6

Source: from ref. 23
[a] Using sodium hydroxide.

Table 13
The Typical Performance of Basket Centrifuges

	Feed			
	Rate (gal/min)	Solids (wt%)	Solids rate (wt%)	Cake solids (wt%)
Basket centrifuge	2–60	2–5	50–95	5–25

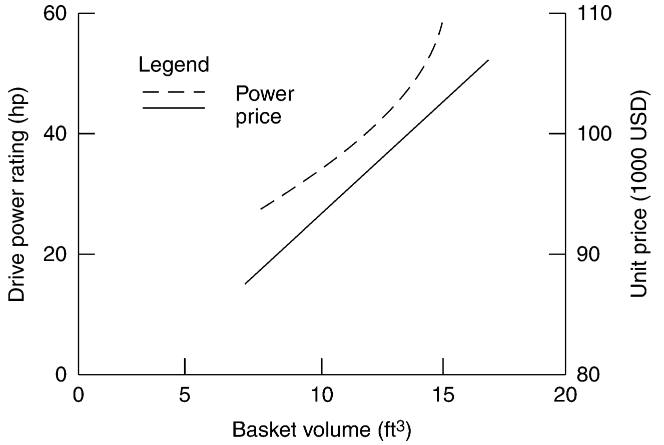

Fig. 10. Unit price and hydraulic drive horsepower for a basket centrifuge (*Source:* US EPA, 1982).

Figure 12 shows the influence of process and centrifuge variables on performance (31). The centrifuge variables (bowl length, bowl diameter, conical bowl angle, and so on.) are fixed during manufacturing. Plant personnel can control pool depth (by adjusting the central weir) and centrifuge speed. Operators can control feed rate and flocculant rate. A detailed description of the process variables and their results are given in reference.

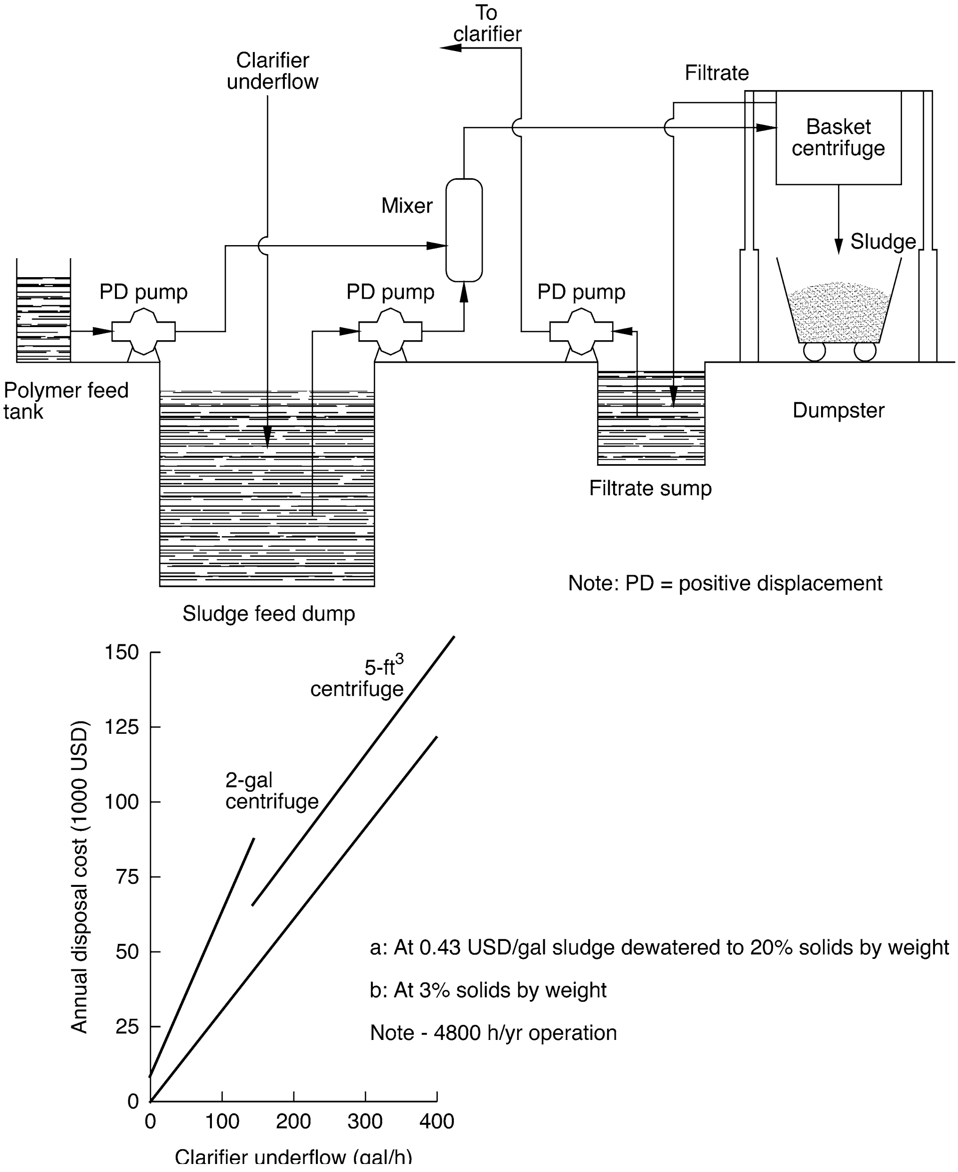

Fig. 11. Basket centrifuges: **(A)** dewatering system with auxiliary equipment **(B)** annual sludge disposal cost (*Source:* US EPA, 1982).

Wakeman (30) described a two-stage centrifuge with a horizontal axis. The bowl has a solid section and a perforated section. A particulate bed forms in the solid section and is moved to the perforated section by a scroll conveyor. Here the solids undergo further dewatering and both solids and liquid are discharged continuously. An oscillating centrifuge (32) can dewater both coarse and fine coals. The cost is 0.03–0.05/USD/t of clean coal. The McNally Wedag horizontal vibrating centrifuge (33) dewaters ultrafines. The vibrations move the solids to the larger diameter of a basket centrifuge and

Table 14
Installed Investment and Annual Cost for Electroplating Sludge Treatment Operation by Using Basket Centrifuges

Feed sludge (gal/h)[a]	Installed investment cost[b]	Annual[c]
50	34,000	37,000
100	34,000	70,000
150	48,000	85,000
200	176,000	89,000
250	176,000	102,000
300	176,000	115,000

Source: from ref. 27.

[a] Initial 3% solids.

[b] An auxiliary equipment.

[c] Operating + fixed + disposal.

Table 15
Typical Centrifuge Data for Coal

Centrifuge type	Particle size	Feed rate (t/h)	Cake moisture (%)	Solids recovery (%)
Horizontal solid bowl (29)	Smaller than 325 mesh	2.5–30	30–35	90–99
Horizontal screen bowl	Larger than 28 mesh	60–100	12–20	90–98
2-Stage (solid and screen) bowl (30)	590 µm × 0	27515	14	96–98
Oscillatory centrifuge (24)				
Coarse coals	1 1/4 × 1/4 in.	Up to 250	2–2.5	97–99
Fine coals	1/4 in. × 28 mesh	Up to 250	6–6.5	97–99
McNally-Wedag vibratory centrifuge (33)				
Coarse coals	$0.5 \times 0.15\ \text{mm}^2$	20	16.8	98
Fine coals	$10 \times 0.5\ \text{mm}^2$	20	7.7	98
Deep-pool horizontal solid bowl design (31)	325 mesh	15–30	3–45	85–99

Data includes effect of polymer addition to improve performance.

also prevent clogging of the basket openings. The effluent water is collected at the other end.

While maintaining centrate quality about 30% increase in throughput can be achieved by optimizing the cylindrical clarifying volume. This is done by redesigning the solid bowl by positioning the conical–cylindrical bowl intersection so as to increase the spillover pool depth. Testing on pilot and full-scale plants with the deep pool design revealed solid feed rates more than 15–30 t/h with 0.1–0.5 gm/kg of polymer producing 55–70% solids cake. Figure 13 shows the effects of pool depth design on cylindrical clarifying volume and "G" level at discharge (31).

Process and centrifuge variables

Centrifuge performance	Polymer dosage	Feed rate	Differential	Pool depth	G'S	% - 325M
Rated centrifuge capacity	→	—	↗	↗	↗	↘
Centrate clarity	↗	↘	↗	↗	↗	↘
Cake percent solids	↘	↗	↘	↘	↗	↘
Polymer* dosage	—	↗	↘	↘	↘	↗

*Polymer dosage required for 100% recovery of the suspended solids

Fig. 12. Process and centrifuge variable influence on centrifuge performance ref. 31.

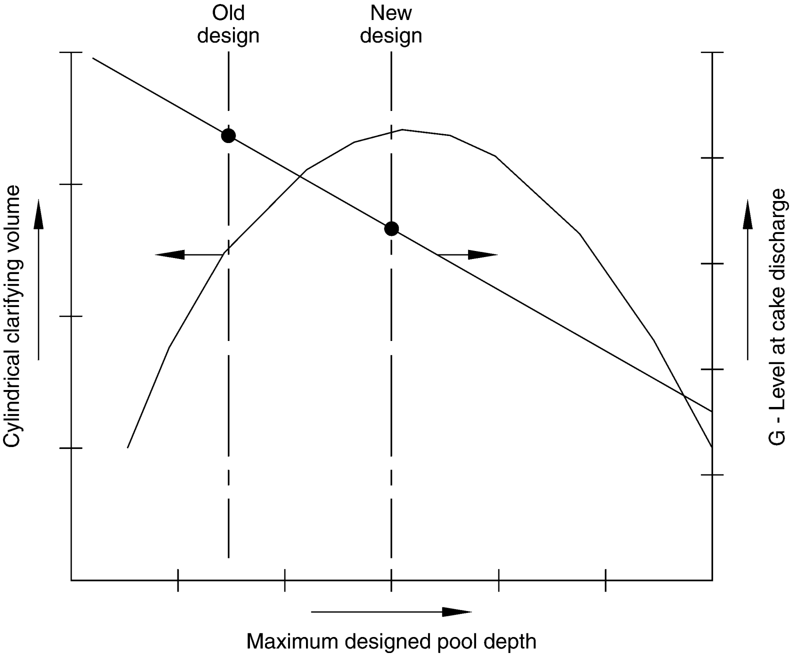

Fig. 13. Effects of pool depth on cylindrical volume and 'G' level at discharge ref. 31.

5.5. Separation of Metallurgical Refinery Sludge by Centrifugation

Sludge dewatering pilot studies at Teledyne Wah Chang Albany evaluated the performance of a solid-bowl centrifuge (34). Differential speed, pond depth and polymer dose were found to influence cake dryness, solids capture and overall centrifuge performance. The abrasiveness of the sludge necessitated the use of a tungsten carbide, hard surface scroll. A feed of 5% solids, gave a cake with 20% solids and a centrate with 0.1% solids for a polymer dosage of 2–5 g/kg and a differential speed of 25 rpm.

5.6. Separation of Cannery Waste Biological Sludge by Centrifugation

Large quantities of biological sludge are generated by treatment of wastewater from fruit processing plants. This sludge can be dewatered for use as cattle feed. A pilot plant scale study was conducted by Washington state, centrifuge of 76 cm diameter was operated at 1300 G. Using the clarifier underflow as centrifuge feed gave 84 g dry solids/cg cakes. It was also found that centrate quality deteriorated as hydraulic loading increased. The centrate quality also deteriorated as the bowl filled with cake. This happened within one solids retention time, irrespective of hydraulic retention time and feed solids concentration.

5.7. Separation of Potato Wastes by Centrifugation

Potato waste is made up of peel, process liquid, wash-down water, volcanic ash, and so on. Approximately 2 MGD (125 t/mo of dry solids) is shipped out of just one plant at Ore-Ida at Boise, Idaho. Recently the plant replaced seven basket centrifuges with three decanter centrifuges, resulting in an annual savings of 65,000 USD. Dewatered waste is primarily sold to animal food processors. A decanter centrifuge is used to achieve the desired solids concentration of 14–15%.

6. CENTRIFUGATION DESIGN CONSIDERATIONS

6.1. General Guidelines for Selecting a Centrifuge for Sludge Dewatering

Many factors affect the performance of centrifuges. Some factors affecting the choice of centrifuges are indicated as follows (37–58,68):

a. Particle size and size distribution.
b. Density of solid.
c. Density of Liquid.
d. Viscosity of liquid.
e. Temperature.
f. pH of sludge.
g. Abrasion properties of sludge.
h. Volatility of solids.
i. Throughput.
j. Final solids concentration.
k. Solids recovery.

It is difficult to consider all the factors during centrifuge selection. A trade-off is probably necessary before assigning a priority to each (39–42). It is recommended that experimental data be obtained on the actual sludge for which a centrifuge is to be selected. Figure 14 can be used as a general guideline for selection of centrifuges. Some points to be considered before a final selection is made are:

a. Levels of solids concentration desired.
b. Amount of flocculant required per pound of dry solids.
c. Suitability of sludge to be dewatered.
d. Comparison with other types of solid–liquid separation devices.
e. Capital, operating, and maintenance cost of equipment.

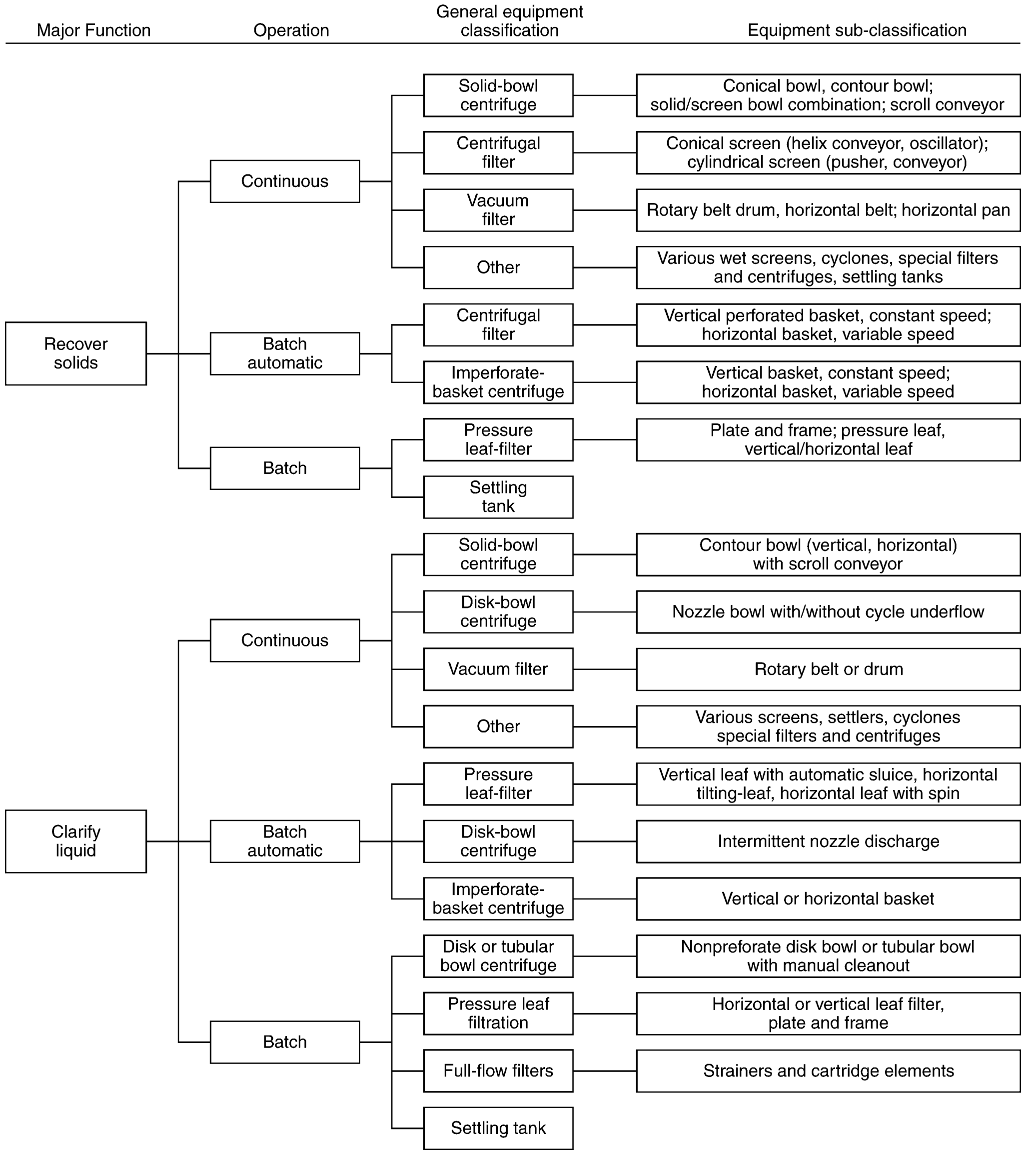

Fig. 14. How to approach a centrifugation problem.

6.2. Centrifuge Manufacturers

A partial list of centrifuge manufacturers are provided as a reference as follows:

a. Alfa Laval—Fort Lee, NJ.
b. Baker Perkins—Saginaw, MI.
c. Bird Machine Co.—South Walpole, MA.
d. Centrico, Incorporated—Englewood, NJ.
e. DeLaval Separator—Poughkeepsie, NY.
f. Dorr Oliver—Stamford, CT.
g. Ingersoll Rand, Eimco Division—Nashua, NH.

h. Komline Snaderson—Peapack, NJ.
i. Pitmar Company—Rochester, NY.
j. Robatel—Pittfield, MA.
k. Sharples, Pennwalt—Philadelphia, PA.
l. Swenson Evaporator—Harvey, IL.
m. Tolhurst Centrifugal Div. Amtec.—East Moline, IL.
n. Wemco Courterbex, Div. Arthur McKee—San Francisco, CA.
o. KHD Humdolt Wedag—Atlanta, GA.
p. Western States—Hamilton, OH.

6.3. Materials for Centrifuge Construction

Stainless steel is typically used for the construction of the three centrifuges as given (1). The bowls for the disc centrifuge are always made of stainless steel. Carbon steel alloys and different coatings are sometimes used for basket centrifuges. Alternative materials for the solid-bowl centrifuge are carbon steel, and titanium is used for special corrosion problems. Covers for disc centrifuges may also be constructed of carbon or cast steel (1).

6.4. Advantages and Disadvantages of Centrifugation in Various Applications

The facts on centrifugation are summarized in this section for design considerations. Centrifugation is a widely used process for concentrating and dewatering sludge for final disposal. The process offers the following advantages:

a. Low capital costs compared with other mechanical equipment.
b. Moderate operating cost, provided flocculants are not required.
c. A totally enclosed unit that minimizes odors.
d. A simple and compact unit.
e. Chemical conditioning of the sludge is often not required.
f. The ability to process a wide variety of solids.
g. Minimum supervision required.

Disadvantages associated with centrifugation are:

a. Without the use of chemicals, solids capture is often poor.
b. Substantial chemical costs.
c. Trash must often be removed from the centrifuge feed by screening.
d. Lower percentage of cake solids than those resulting from vacuum filtration.
e. Higher maintenance costs than vacuum filtration.
f. Fine solids (in concentrate) that escape the centrifuge may resist settling when recycled to the head of the treatment plant and gradually increase in concentration, eventually raising effluent solids level.

Centrifuges applicable to sludge thickening and dewatering fall into three general classifications: disc, basket, and the currently popular solid-bowl. Basically, centrifuges separate solids from liquids through sedimentation and centrifugal force. Process variables in centrifugation include feed rates, sludge solids characteristics, feed consistency, and chemical additives. Machine variables include bowl design, bowl speed, pool volume, and conveyor speed.

The main objectives in centrifuge design are cake dryness and solids recovery. The effects of various parameters on these two factors are summarized in Table 16.

Table 16
Summary of the Effect of Various Parameters on Centrifuge Performance

To increase cake dryness	To increase solids recovery
Increase bowl speed	Increase bowl speed
Decrease pool volume	Increase pool volume
Decrease conveyor speed	Decrease conveyor speed
Increase feed rate	Decrease feed rate
Decrease feed consistency	Increase temperature
Increase temperature	Use flocculants
Do not use flocculants	Increase feed consistency

Operating data reported in the literature indicate that raw primary and digested primary sludge dewater easily. With the addition of polymer, a centrifuge (Fig. 2) can produce 25–40% cake solids with better than 90% recovery.

However, waste-activated sludge, is difficult to thicken or dewater with centrifugation. High polymer dosage will be required to produce 8–10% cake solids and 90% recovery.

6.5. Design Criteria, Input Data, and Design Parameters

Design criteria for centrifugation systems are scarce. One criterion used in determining the size of centrifuge required is the power requirement per gallon per minute of inflow (0.5–2 hp/gpm). Generally, a power requirement of 1 hp/gpm of inflow is applicable to most commercially manufactured centrifuges (42–59).

The input data for centrifuge design includes:

a. Sludge flow (gpd).
b. Sludge concentration (%).

Design parameters include at least the following:

a. Centrifuge power requirement (hp/gpm = approx 1 hp/gpm).
b. Operation hours per day (h).
c. Operation days per week (d).
d. Number of units.
e. Excess capacity factor (1.25).
f. Chemical dosage by dry weight of solids (%).

6.6. Design Procedure

6.6.1. Calculation of Total Power Requirement

The total power required for centrifuge operation can be calculated by the following equation:

$$\mathrm{THP} = \frac{(\mathrm{SF})\,(\mathrm{CPR})\,(7\ \mathrm{d/wk})\,(\mathrm{ECF})\,(24\ \mathrm{h/d})}{(1440\ \mathrm{min/d})\,(\mathrm{d/wk})\,(\mathrm{h/d})} \tag{1}$$

where THP is the total power required (hp); SF is the sludge flow (gpd); CPR is the centrifuge power requirement (hp/gpm [approx 1]); DPW is the operation/wk (d/wk); ECF is the excess capacity factor (1.25); and HPD is the operation/d (h/d).

6.6.2. Calculation of Power Per Each Centrifuge Unit

Always use duplicate units. The power per centrifuge unit can be calculated by the following equation:

$$\mathrm{HPU} = \frac{\mathrm{THP}}{\mathrm{NU}} \tag{2}$$

where HPU is the horse power per unit (hp/unit); THP is the total power required (hp); and NU is the number of centrifuge units.

6.6.3. Selection of Centrifuge

Centrifuge hp should be compared with the manufacturer's specification when selecting a centrifuge that meets the requirements.

6.6.4. Calculation of Chemical Requirements

The chemical requirements of centrifugation units can be calculated by the following equations:

$$\mathrm{CR} = \frac{(\mathrm{SF})\ (7\ \mathrm{d/wk})\ (C_d)\ (8.34\ \mathrm{lb/gal})\ (100 - C_i)}{(\mathrm{h/d})\ (\mathrm{d/wk})\ (100)\ (100)} \tag{3}$$

where CR is the chemical requirements (lb/h); SF is the sludge flow (gpd); C_d is the chemical dosage, percent of dry weight of solids fed to unit; HPD is the daily operation (h/d); DPW is the weekly operation (d/wk); and C_i is the initial moisture content of sludge (%).

6.6.5. Design Output Data

The output data from centrifugation system design will include the following:

a. Power required/unit.
b. Number of units.
c. Chemical requirements (lb/h).
d. Sludge flow (gal/d).
e. Initial solid concentration (%).
f. Daily operation (h/d).
g. Weekly operation (d/wk).

7. OPERATION AND MAINTENANCE

7.1. Troubleshooting

The Water Pollution Control Federation (WPCF) Technical Practice Committee (4) provides troubleshooting guidelines to address the typical operations problems of centrifuges. Table 17 summarizes the WPCF recommendations.

7.2. Preventive Maintenance

Preventive maintenance for the solid bowl unit is discussed later. The occurrence and amount of inspection depends on the type of sludge solids dewatered and maintenance service circumstances. If the sludge is abrasive, the first inspection is recommended after 1000 h of use. If the sludge is not abrasive and the machine has a low speed, the first inspection is recommended after 3000–4000 h. Extensive maintenance normally requires disassembly of major parts. High-speed balancing is needed and a manufacturer must be contacted for high-speed equipment operating at a rate more than

Table 17
Troubleshooting Guide for Centrifugal Thickening

Problem	Probable cause	Solutions
Inconsistent thickened solids or centrate	Varying inflow feed rate	Install flow monitoring control system
	Varying inflow concentrate	Monitor and control feed sludge
	Improper machine adjustment	Make appropriate centrifuge adjustments
	Varying chemical feeds	Install flow monitoring control system or verify chemical batch consistency
Improper solids capture	Improper machine settings	Adjust machine
	Improper chemical feed rate	Monitor and adjust chemical feed rate
	Low solids in sludge inflow	Monitor sludge inflow characteristics and adjust sludge feed rate or centrifuge running time
High bearing temperature, high vibration, or high motor amperage	Internal machine problems, such as low oil to bearings	Shut down machine and consult manufacturer
Excessive vibration	Machine may not have been adequately flushed and cleaned from last usage	Flush machine thoroughly if the shut-down had been prolonged, it may require dissembling and manual cleaning
Foaming centrate	Excessive polymer usage	Reduce polymer and adjust dilution meter
Increased or changed sound	Change in feed rate	Check and adjust feed rate as necessary
	Concentration equipment malfunction	Determine source of noise and repair

600–1200 G. Otherwise, the repair is easier and can take place in the field. The typical schedule maintenance is comprised of periodic lubrication and flushing the centrifuge bowl every shift at shutdown. Bushings, bearings, and seals should be inspected regularly and replaced when needed. Plant workers should have spare parts on hand to replace shear pins, main bearings, seals, and conveyor bushings (45). The following should be checked periodically to prevent early breakdowns:

a. Amount of oil in the reservoir.
b. Flow of oil to the bearings.
c. Flow of cooling water and the oil temperature to ensure operations in the correct range.
d. Machine vibration, which may result in damaged bearings and other parts.
e. Ammeter reading for proper centrifuge loading.
f. The bearing for uncommon noises.
g. Overheated bearings by touching them. Hot bearings may be over or under lubricated.
h. Leaks.

7.3. Noise and Odor Control

The solid-bowl centrifuge is totally enclosed, thus noise is controlled and odors are restricted. To function properly centrifuges require ventilation; therefore, if venting causes an odor problem, the released gas should be treated. Compared to other mechanical dewatering systems such as vacuum filters, filter belt presses, and filter presses, the contained centrifuge gas is easier to collect and treat and has a less significant effect on building air quality. Because of this, the chances of harmful odor being released from the dewatering room ventilation are minimal (44).

Usually the use of a centrifuge will require odor control because of the openings at the cake and centrate drop chutes. Although these openings are small, odors that seep out will be strongly concentrated because of the containment cover. Optimizing upstream operations and adding chemicals can reduce these odors. Odors from centrifuge operation are a function of influent feed, which is why optimization of upstream operation will reduce odors. The addition of chemical oxidants like potassium permanganate and hydrogen peroxide before centrifugation will also reduce the amount of odors released (44,68). The aforementioned mechanical dewatering equipments other than centrifuges are uncovered and open to the atmosphere. Therefore, centrifugation is a more suitable alternative for controlling fugitive odor release.

8. DESIGN AND PRACTICAL APPLICATION EXAMPLES

8.1. Example 1: Centrifugation System Design

Calculate the total power required for a centrifugation system and the hp per centrifuge unit. Then select the commercially available centrifuge units, all based on the following given information and equation.

$$\mathrm{THP} = \frac{(\mathrm{SF})(\mathrm{CPR})\ (7\ \mathrm{d/wk})\ (\mathrm{ECF})\ (24\ \mathrm{h/d})}{(1440\ \mathrm{min/d})\ (\mathrm{d/wk})\ (\mathrm{h/d})} \tag{1}$$

where THP is the total power required (hp); SF is the sludge flow (7000 gpd); CPR is the centrifuge power requirement (1 hp/gpm); DPW is the weekly operation (5 d/wk); ECF is the excess capacity factor (1.25); and HPD is the daily operation (16 h/d).

Solution

The first step is to calculate the total power required for entire centrifugation system.

$$\mathrm{THP} = \frac{7000\ (1)\ (1.25)\ (24)}{(1440)\ (5)\ (16)} = 12.8\ \mathrm{hp}$$

The next step is to determine the hp of each centrifuge.

$$\mathrm{HPU} = \frac{\mathrm{THP}}{\mathrm{NU}} \tag{2}$$

where HPU is the horse power per unit (hp/unit); THP is the total power required (12.8 hp); and NU is the number of units = 2 units, then

$$\mathrm{HPU} = \frac{12.8}{2} = 6.4\ \mathrm{hp}$$

The third step is to compare hp with the manufacturer's specification and select a centrifuge that meets the power requirement.

8.2. Example 2: Centrifugation System Chemical Requirements

Calculate the chemical requirements of a centrifugation system based on the following given information and design equation:

$$\mathrm{CR} = \frac{(\mathrm{SF})\ (7\ \mathrm{d/wk})\ (C_d)\ (8.34\ \mathrm{lb/gal})\ (100 - C_i)}{(\mathrm{h/d})\ (\mathrm{d/wk})\ (100)\ (100)} \tag{3}$$

where CR is the chemical requirements (lb/h); SF is the sludge flow (7000 gpd); C_d is the chemical dosage, percent of dry weight of solids fed to unit (1%); HPD is the daily operation (16 h/d); DPW is the weekly operation (5 d/wk); and C_i is the initial moisture content of sludge (97.5%).

Solution

$$\mathrm{CR} = \frac{(7000)\ (7)\ (1)\ (8.34)\ (100 - 97.5)}{(16)\ (5)\ (100)\ (100)} = 1.28\ \mathrm{lb/h}$$

8.3. Example 3: Centrifugation System Cost Estimation

The cost data presented in Table 14 and Figs. 6–11 all date back to 1882. How can the old cost data be updated to the present cost?

Solution

If equipments cost must be escalated to the current year, the chemical engineering (CE) equipment index (Table 18) or equivalent can be used. Monthly indices for 5 yr are provided in Table 18. The following equations can be used for converting the past cost to the future cost, or vice-versa.

$$\mathrm{Cost}_b = \frac{\mathrm{Cost}_a \times \mathrm{Index}_b}{\mathrm{Index}_a} \tag{4}$$

where Cost_a is the cost in the month to year of *a* (USD); Cost_b is the cost in the month-year of *b* (USD); Index_a is the CE Fabricated Equipment Cost Index in the month-year of *a*; and Index_b is the CE Fabricated Equipment Cost Index in the month-year of *b*.

Although it should be noted that the CE Fabricated Equipment Cost Indices (60,62,64) are recommended for Index_a and Index_b, the ENR Cost Indices (61,63), and the US Army Cost Index (65) can also be adopted for updating the cost. Cost data for construction and operation and maintenance (O&M) originate from a variety of reference sources and reflect different time periods and geographic locations. In addition, regional factors must also be considered in cost estimation. Readers are referred to another book for details about environmental equipment cost estimation (4).

8.4. Example 4: Centrifugation Case Study

A plant in Chicago West Southwest plant in Stickney, Illinois, is a waste-activated conventional aeration facility that has a combined storm and wastewater average flow rate of 34.8 $\mathrm{m^3/s}$ (800 MGD) and a maximum flow rate of 52.2 $\mathrm{m^3/s}$ (1200 MGD). The sludge contains a combination of waste from primary and waste-activated sludge

Table 18
CE Fabricated Equipment Index

Date	Index
April 2000	437.4
February 1990	389
January 1990	388.8
December 1989	390.9
November 1989	391.8
October 1989	392.6
September 1989	392.1
August 1989	392.4
July 1989	392.8
June 1989	392.4
May 1989	391.9
April 1989	391
March 1989	390.7
February 1989	387.7
January 1989	386
December 1988	383.2
November 1988	380.7
October 1988	379.6
September 1988	379.5
August 1988	376.3
July 1988	374.2
June 1988	371.6
May 1988	369.5
April 1988	369.4
March 1988	364
February 1988	363.7
January 1988	362.8
December 1987	357.2
November 1987	353.8
October 1987	352.2
September 1987	343.8
August 1987	344.7
July 1987	343.9
June 1987	340.4
May 1987	340
April 1987	338.3
March 1987	337.9
February 1987	336.9
January 1987	336
December 1986	335.7
November 1986	335.6
October 1986	335.8
September 1986	336.6
August 1986	334.6
July 1986	334.6

(Continued)

Table 18 ***(Continued)***

Date	Index
June 1986	333.4
May 1986	334.2
April 1986	334.4
March 1986	336.9
February 1986	338.1
January 1986	345.3
December 1985	348.1
November 1985	347.5
October 1985	347.5
September 1985	347.2
August 1985	346.7
July 1985	347.2
June 1985	347
May 1985	347.6
April 1985	347.6
March 1985	346.9
February 1985	346.8
January 1985	346.5
December 1984	346
December 1979	273.7
December 1977	226.2

Chemical Engineering, refs. 60 and 62.

flow streams. The first centrifuges were replaced in 1989 and in 1990 and 12 high solids dewatering centrifuges were installed. Nine or ten centrifuges are normally operating 24 h/d, 365 d/yr. The remaining two or three units have either been fixed or are on reserve. Centrifuge capacity is 12.7 L/s (200 gpm) per machine and the resulting cake amount per machine is 30–40 dry t (43).

Three sludge disposal alternatives are utilized, depending on the time of year—landfills, dry beds, or lagoons. The new centrifuges produce a cake solid that is more than 10% that from the previous machines (43).

NOMENCLATURE

THP	Total power required (hp)
C_d	Chemical dosage, percent of dry weight of solids fed to unit
C_i	Initial moisture content of sludge (%)
Cost_a	The cost in the month to year of *a* (USD)
Cost_b	The cost in the month to year of *b* (USD)
CPR	Centrifuge power requirement, (hp/gpm) = 1 hp/gpm
CR	Chemical requirements (lb/h)
DPW	Weekly operation/wk (d/wk)
ECF	Excess capacity factor (1.25)
HPD	Daily operation/d (h/d)
HPU	Horse power/unit (hp/unit)

Index_a	The CE Fabricated Equipment Cost Index in the month-year of *a*
Index_b	The CE Fabricated Equipment Cost Index in the month-year of *b*
NU	Number of units
SF	Sludge flow (gpd)
THP	Total power required (hp)
US EPA	US Environmental Protection Agency
WEF	Water Environment Federation
WPCF	Water Pollution Control Federation

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5

Anaerobic Digestion

Jerry R. Taricska, David A. Long, J. Paul Chen, Yung-Tse Hung, and Shuai-Wen Zou

CONTENTS

1. INTRODUCTION

Conversion of the organic material from 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, 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. Generally the process is considered to be a two-stage biological process involving waste conversion and stabilization (Fig. 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 sludges has long been used to stabilize organic wastes prior to final disposal of these wastes. Among the benefits to be realized from such treatment are:

a. A reduction in the organic content of the sludge.
b. Improved sludge dewaterability.

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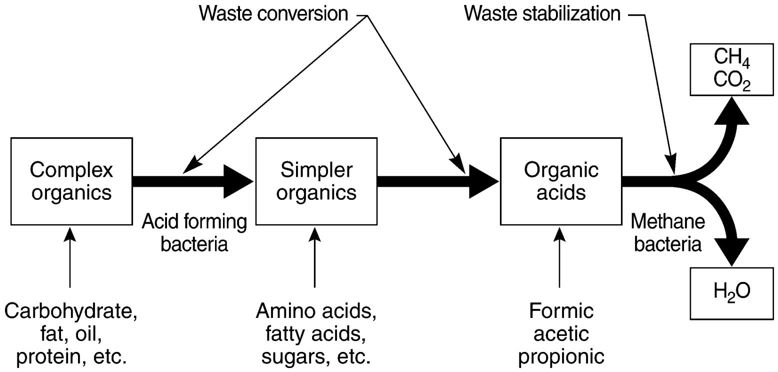


Fig. 1. General anaerobic biological reactions.

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 is produced.
c. No oxygen is required.
d. Nutrient requirements are low (1).

Not only an interest in the anaerobic processes is being generated because of their waste treatment potential, but because in our increasingly energy conscious society, the potential for generating methane from waste materials takes an 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 prior to digestion is necessary. The solid wastes should be first sorted and shredded to a size that will not interfere with the proper functioning of the digester equipment and transport systems. Normally it is necessary to add moisture and nutrients (pH adjustment may be required) to form a slurry that can be heated, prior to feeding the mixture into the digester. Sewage sludge is often 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 the 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 the primary settling of raw wastes can be made by the 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 is available on the raw wastewater and primary clarifier effluent suspended solids concentrations together with an 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 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, in order 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 indicated earlier, the anaerobic digestion process is considered to be a minimum of 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 is generally 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, in order to avoid oxygen toxicity. In order for the complex organics to be utilized by the acid-forming bacteria, they must be first 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 next, Eq. (2) is generally considered to represent the more important reaction.

$$CO_2 + 8H^+ \xrightarrow{\text{Methane-forming bacteria}} CH_4 + 2H_2O \tag{1}$$

$$CH_3COOH \xrightarrow{\text{Methane-forming bacteria}} CH_4 + CO_2 \tag{2}$$

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 utilized. 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 utilization makes it possible to show waste stabilization efficiency for the anaerobic process which 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, which then require a group of methane organisms that are not present in sufficient numbers for a balanced condition to exist. The necessary balance should re-establish itself, unless the influent waste or sludge characteristics or other factors are constantly changing. Figure 2 summarizes the major metabolic pathways involved in the anaerobic digestion process.

2.3. Reactor Configurations

There are four basic reactor configurations that have been used for the design of anaerobic unit processes. The four are as follows:

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 are normally used for the digestion of solid wastes and/or wastewater sludge. Figure 3 illustrates the basic reactor types for the 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 sludge concentration to be accomplished in a single tank.

High-rate digestion systems are designed usually as two-stage systems wherein the primary stabilization of the sludge is accomplished-in the first stage, using a mixed digester. Supernatant separation and sludge concentration, as well as gas storage is provided for the second-stage unmixed digester. The term “stages” as used in conventional engineering practice, does not refer to the biochemical stages of acid production and

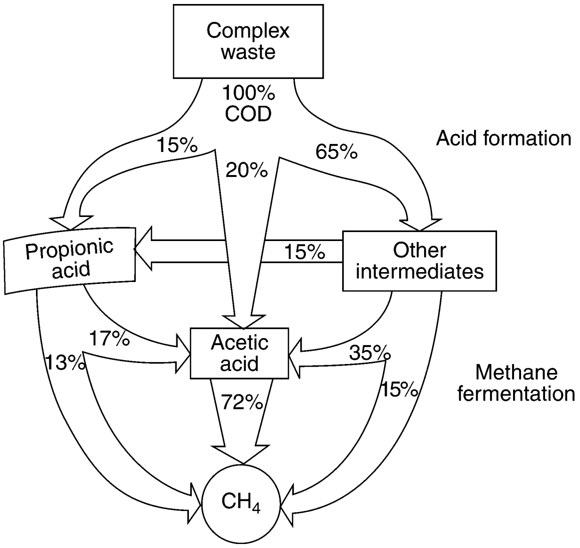
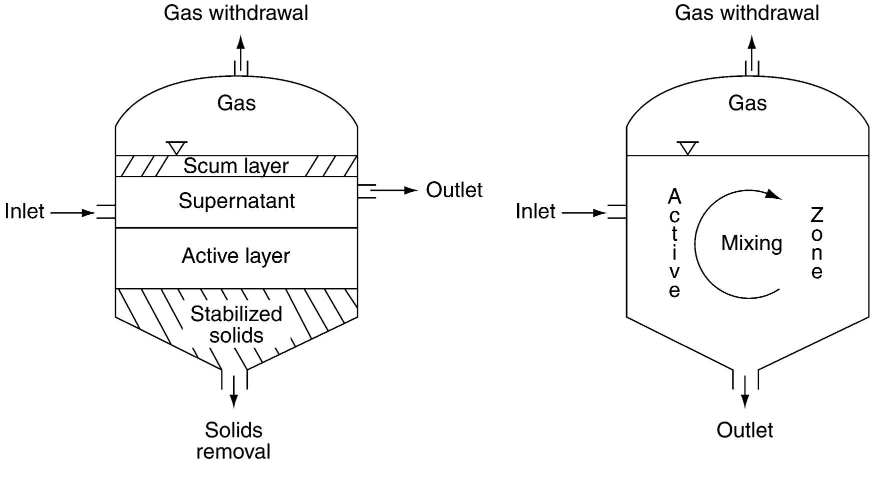


Fig. 2. Metabolic pathways in complex waste stabilization by anaerobic processes.



Conventional High rate

Fig. 3. Digester reactor configuration.

methane formation discussed earlier, 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.

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 (Fig. 4) and the anaerobic filter have been developed (6,7). The anaerobic contact process is simply an anaerobic activated sludge process and is designed

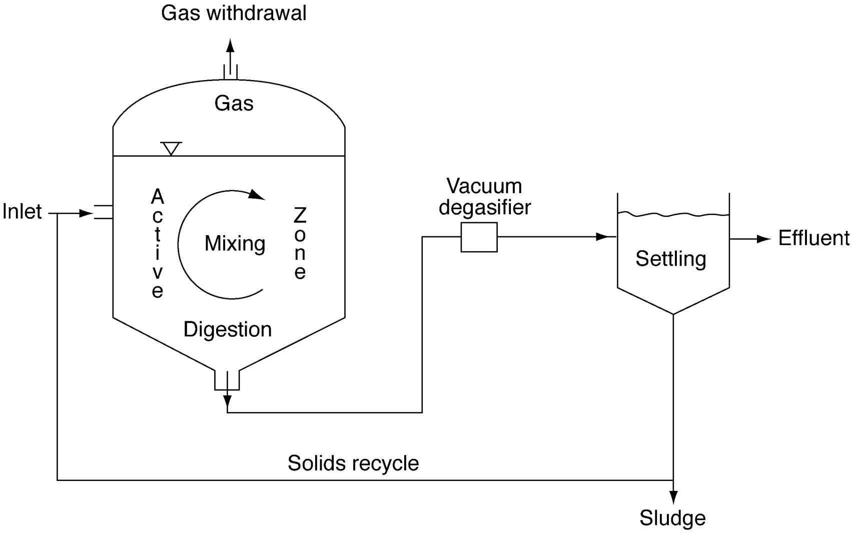


Fig. 4. Anaerobic contact process schematic.

and operated in much the same manner as complete mixed activated sludge 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 the 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 with 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 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 the 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, i.e., kilograms of VS per unit time, normally 1 d/m^3 of the digester capacity.

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

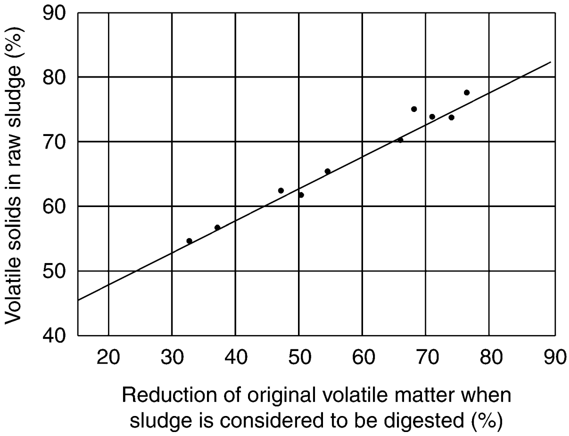


Fig. 5. Reduction in volatile matter by digestion.

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 0.48–1.6 kg VS/d/m^3 (0.03–0.10 lb VS/d/ft^3), whereas the high-rate digester (mixed) normally would be loaded in the range of 1.6–6.4 kg VS/d/m^3 (0.10–0.40 lb VS/d/ft^3). Design loadings will be discussed in more detail in the section on reactor design and sizing below. The degree of stabilization of sludges by anaerobic digestion is best expressed in terms of the VS reduction that has occurred during digestion. Figure 5 illustrates accepted practice in determining when the 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 0.64–1.6 kg VS/d/m^3 (0.04–0.10 lb VS/d/ft^3) should be satisfactory. VS reduction in such mixtures ranges 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 utilized for sludge digestion in current practice. Mesophilic digestion at temperatures ranging from 30 to 37.7°C (86–100°F) is most commonly employed with the majority of installations operating at approx 35°C (95°F). Thermophilic digestion at temperatures ranging from 49 to 57°C (120–135°F) has been used on a limited basis.

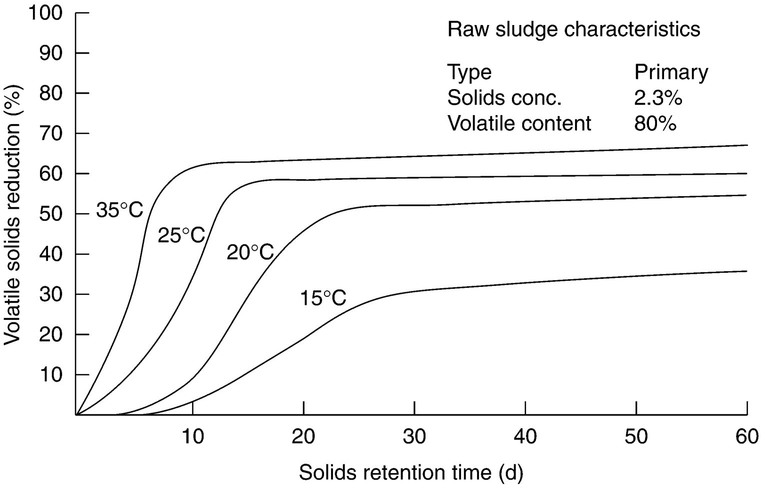


Fig. 6. Effect of solids retention time and temperature on volatile solids reduction in a laboratory-scale anaerobic digester (17).

Generally, biological reactions follow the Arrhenius principle of chemical reaction rates. Figure 6 shows the general relationship between the temperature and digestion efficiency, based on solids retention time (SRT) in days. As can be seen from Fig. 6, digestion at elevated temperatures in the thermophilic range is more efficient in terms of the biological reactions involved (42). 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 the temperature and of mixing, for reasons that have been stated above. In addition to these parameters, VS reduction (waste stabilization) has been shown to be related to the raw sludge characteristics, as shown in Figs. 7 and 8. Conventional digesters (standard rate) normally are designed with detention times of 30–60 d, whereas high-rate systems are designed with detention times of approx 15 d in the first-stage mixed digester (43).

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 the treatment of municipal wastes, but they may not be present in the required amounts in industrial wastes. Industrial waste analyses should include these two parameters. Nutrient requirements normally are approx 11 and 2% of the biological solids weight produced for nitrogen and phosphorus, respectively (1). Pilot studies can be helpful in assessing nutrient requirements for a particular system.

2.7. Gas Production and Utilization

The principal gases produced during the anaerobic digestion process are methane and carbon dioxide. Small amounts of hydrogen sulfide (H_2S) are also normally present and

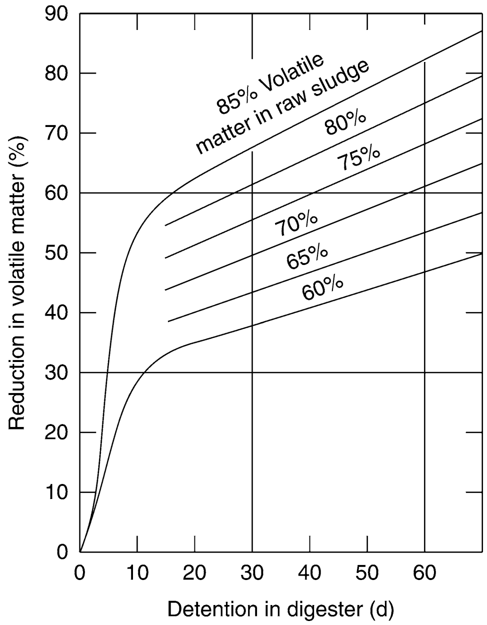


Fig. 7. Reduction in VS in raw sludge (17).

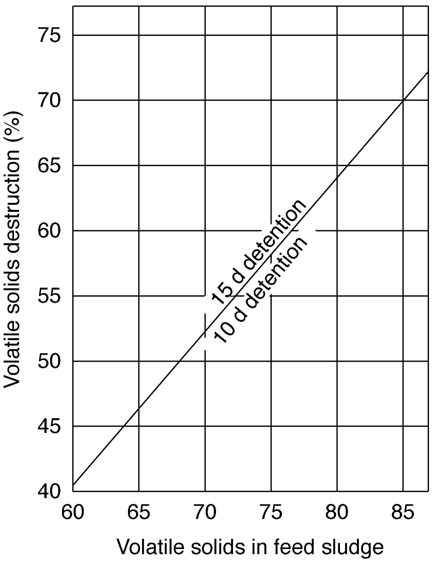


Fig. 8. Expected VS destruction during high-rate digestion (17).

it may be noticeable in terms of the odor characteristics of the digester gas. Gases from a properly operating-digester should contain approx 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 the digester gas is approx 5850 kg-cal/m^3 (656 Btu/ft^3), although this value will vary depending upon the composition of a particular digester gas.

The amount of gas to be expected conventionally has been expressed in terms of: (a) volume of gas produced per unit weight of VS 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) 0.5–0.75 m^3/kg (8–12 ft^3/lb) of VS added, or (b) 0.75–1.1 m^3/kg (12–18 ft^3/lb) of VS 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 utilization, and the amount of biological solids produced are known. Equation (3) expresses this relationship in metric units.

$$C = 0.35\,(eF - 1.42\,A) \tag{3}$$

where C is the methane produced, m^3/d; e is the efficiency of waste utilization; F is the BOD_5 added, kg/d; and A is the VS produced kg/d.

Digester gases have been utilized 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 in order 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 the industrial wastes. In the second case, the treatability study will define the 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 by using either the batch or continuous flow reactors. The batch system is less complex to setup and operate and is more applicable to the evaluation of sludges rather than for the treatment of total waste streams. Continuous flow reactors with some form of solids recycle is advantageous for evaluating anaerobic contact treatment. A schematic diagram of a batch system reactor setup for anaerobic treatability studies is shown in Fig. 9.

For sludge digestion studies, the operating variables to be investigated are detention time and VS 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 SRT can be considered to be the same. Sludge additions can be made once a day, but

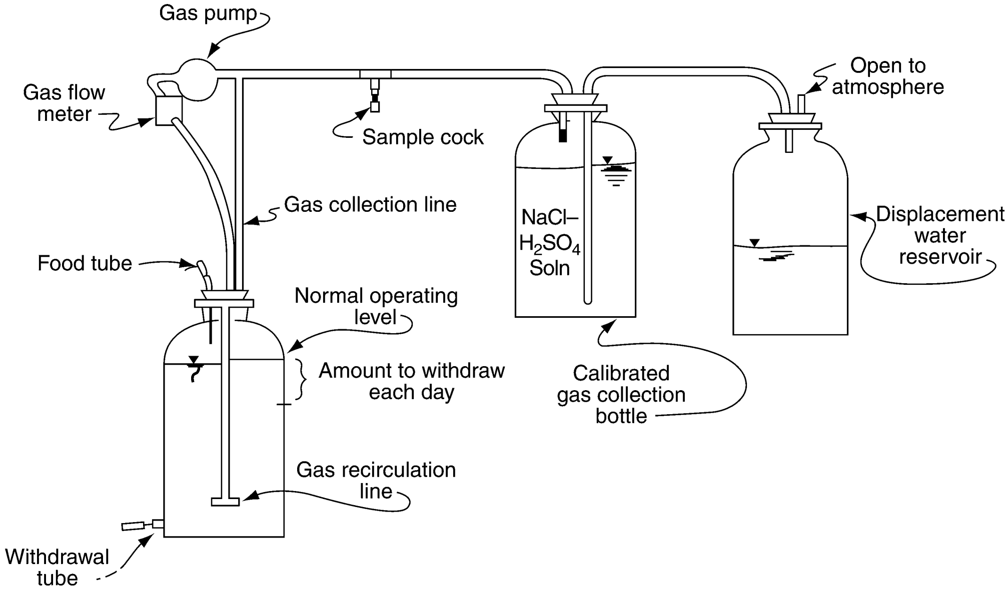


Fig. 9. Pilot digester setup for anaerobic process.

twice daily is preferable. If possible, digester feeding schedules to be used in the full-scale installation should be simulated. VS 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 procedure is as follows:

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 VS concentration for the VS loading to be used at a particular detention time. Sludge can be made up ahead and stored at 4°C for up to 1 wk.
6. Digesters can be fed once or twice daily; and 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 the 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 (<5%) in most instances.

8. Analyses to be performed include the following:
   a. Total and VS 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 the data should include as follows:
   a. VS loading and reduction under the various operating conditions.
   b. Gas production per unit weight of the VS added and destroyed.
   c. Plots of VS loading, VS reduction, volatile acids, gas production vs time for each unit.

If the study is for the purpose of evaluating the anaerobic contact process for the treatment of the total waste stream rather than for evaluation of sludge digestion, essentially the same procedure as outlined earlier 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 removing 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 3 times/wk, in order to evaluate various loadings and removal efficiencies, solids production, kinetic coefficients, and so on; similar to the procedures used in the 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 VS loading, temperature, extent of mixing, and so on. In the absence of calculations that justify the basis of design, the GLUMRB Standards (9) 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 of up to 80 lb of VS per 1000 ft^3 of volume/d (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 the sludge through an external heat exchanger, the system may be loaded at a rate of up to 40 lb of VS per 1000 ft^3 of volume/d (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 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 $VS/m^3/d$ (0.10–0.40 lb $VS/ft^3/d$).

Table 1 shows VS loading factors as a function of the sludge concentration and the hydraulic retention time in the digester. These loading factors can be used to size high

Table 1
Effect of Sludge Concentration and Hydraulic Retention 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] From ref. 10.

HRT = Hydraulic Retention Time.

rate digesters. If mixed organic and chemical sludges are to be digested, the volume of the digester must be increased over that calculated time, 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 VS in the sludge to be digested to the 75% VS sludge used as the basis for calculating the tabular values.

Reactor design based on the 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 the anaerobic systems in this regard. Only a 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 the 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 earlier.

Development of mathematical models for describing the biological wastewater treatment processes has been described elsewhere (10,11). Lawrence (12) has presented the development of models for the application of process kinetics to the design of anaerobic processes. The working relationships for these models are summarized in Table 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 3 for the design. Examples of the use of the models will be presented in a later section.

As anaerobic systems are low growth systems, mean cell residence times or solids retention times (θ_c) must be long enough to avoid washout of the active microbial mass. Solids recycle in the anaerobic contact process helps to maintain the necessary

Table 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	$E_s = \dfrac{100(S_o - S_1)}{S_o}$	(4)	$E_s = \dfrac{100(S_o - S_1)}{S_o}$	(4)
Effluent waste concentration	$S_1 = \dfrac{UK_s}{k - U}$	(5)	$S_1 = \dfrac{UK_s}{k - U}$	(5)
Microorganism concentration in reactor	$X = \dfrac{Y(S_o - S_1)}{1 + k_d\theta_c}$	(6)	$X = \dfrac{Y(S_o - S_1)\theta_c}{1 + k_d\theta_c\theta}$	(7)
Excess microorganism production rate	$A = \dfrac{YQ(S_o - S_1)}{1 + k_d\theta_c}$	(8)	$A = \dfrac{YQ(S_o - S_1)}{1 + k_d\theta_c}$	(8)
Hydraulic retention time (V/Q)	$\theta = \theta_c$	(9)	$\theta = \theta_c[1 + r - r\,(X_r/X)]$	(10)
Solids retention times: General	$\dfrac{1}{\theta_c} = \dfrac{YkS_1}{K_s + S_1} - k_d$	(11)	$\dfrac{1}{\theta_c} = \dfrac{YkS_1}{K_s + S_1} - k_d$	(11)
Solids retention times: Limiting	$\dfrac{1}{\theta_c^m} = Yk - k_d,$ $S_o >> K_s$	(12)	$\dfrac{1}{\theta_c^m} = Yk - k_d,\ S_o >> K_s$	(12)

[a] Nomenclature for the kinetic models is listed at the end of this chapter.
[b] From refs. 10,11.

Table 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] From ref. 10.

active biomass in the reactor, while maintaining a desired short hydraulic retention time (θ).

Design parameters for anaerobic filters are not well-defined in the current practice. Pilot studies should be conducted to determine the organic and hydraulic loading rates, and so on.

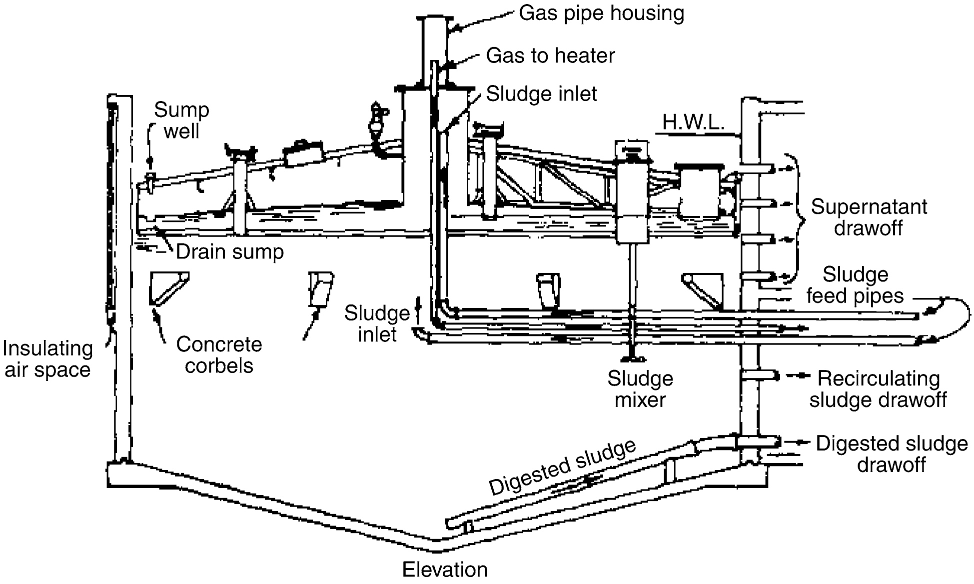


Fig. 10. Typical digester section.

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 (Fig. 10). Tank diameters vary from approx 6 to 34 m (20–111.5 ft) in increments to accommodate standard digester covers. Vertical side wall depths normally range from approx 6 to 12 m (20–39.4 ft). The bottom slope should be at least 1 vertical to 4 horizontal when the sludge is removed by gravity and reduced to 1 vertical to 12 horizontal when the sludge is removed with suction mechanism (9).

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 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 that can be critical during periods when digesters are under stress or when mechanical breakdown 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 earlier. As indicated earlier, vacuum degasification (approx 5 cm Hg of vacuum) of the digester effluent is usually necessary in the anaerobic contact system, in order to achieve the good solids separation that is essential for the recycle of solids. Design of the sludge separator is critical. Unfortunately, few data are available to assist the designer in selecting appropriate design criteria. Settling column tests of the degasified effluent

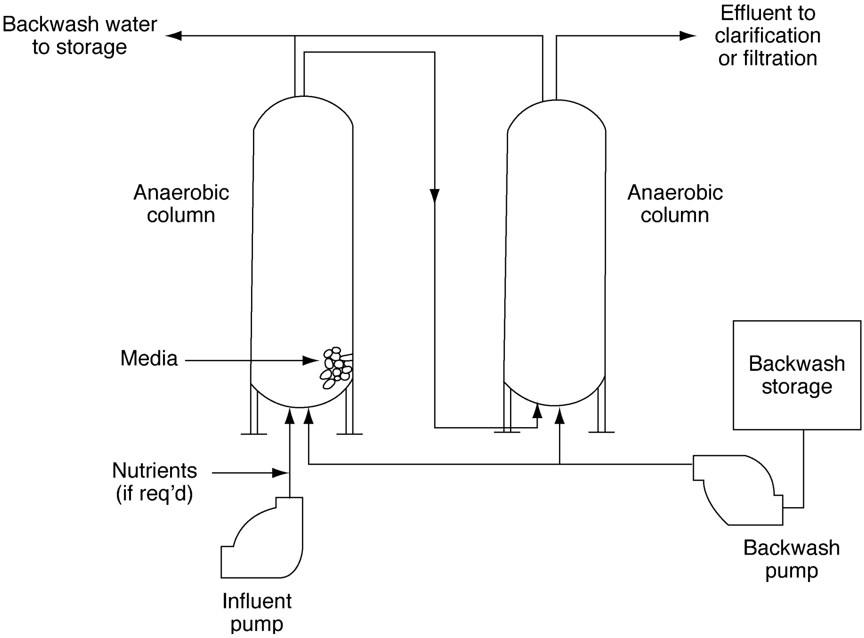


Fig. 11. Process schematic for anaerobic contact column.

should provide guidelines for selecting overflow rates and detention times for the full-scale installation. The use of the solids flux approach as described by Dick and Young (13) should be investigated for applicability to the anaerobic contact process. As the typical sludge generated in this process is a low density flocculent sludge, conventional plow-type clarifier equipment is not suitable for the 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 are normally used with the anaerobic filter system. The particular system configuration to be used will vary with the specific requirements of a particular installation. Figure 11 shows a typical process schematic for the anaerobic filter process. Various synthetic media are available for the 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 the use in anaerobic digestion systems. It is advisable to consult with them, regarding the application of their equipment for a particular installation.

3.4.1. Digester Covers

Digester covers are normally one of the 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

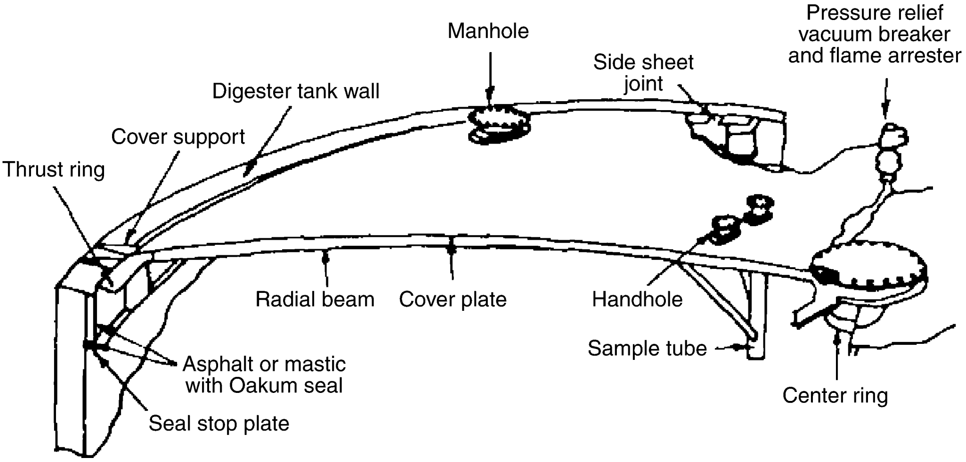


Fig. 12. Fixed digester cover and appurtenances.

or steel, whereas floating covers are usually 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 (5–15% 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 the 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, 61–72 cm (24–28 in.) openings, should be provided in the digester cover. Gastight, quick-opening sample tubes also should be provided. Special precautions are necessary to prevent fire or explosions, whenever digester covers are opened. Figures 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 for 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 the samples taken from various locations in the digester should not vary more than 10% in suspended solids concentration. Other recommendations suggest that at least 3 turnovers/d of the entire digester contents be provided in the recirculation system.

Gas recirculation systems have proven to be very popular in current practice. Several manufacturers furnish equipment that utilizes 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 15 shows a layout of

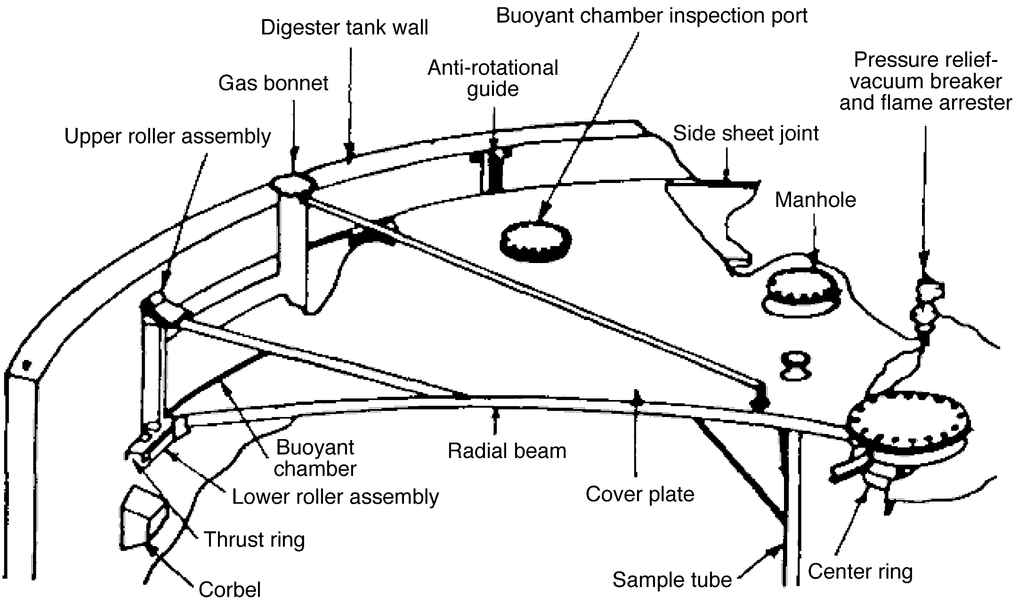


Fig. 13. Floating digester cover and appurtenances.

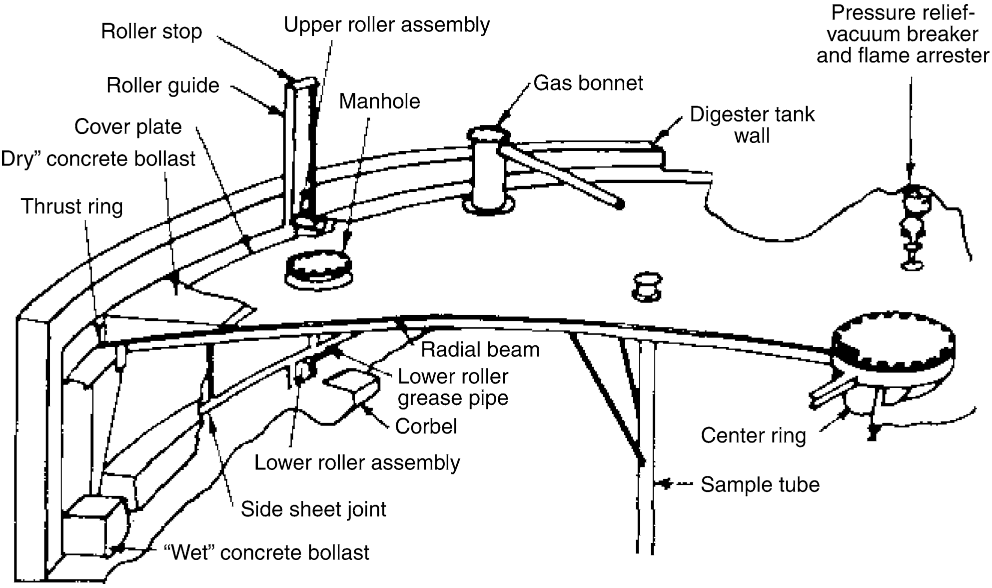


Fig. 14. Digester gas holder cover and appurtenances.

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 16 shows an example of one type of draft tube mixer. Mixing can also be provided by means of turbine-type mixers as shown in Fig. 17. Details on any of the mixing systems are available from the manufacturers.

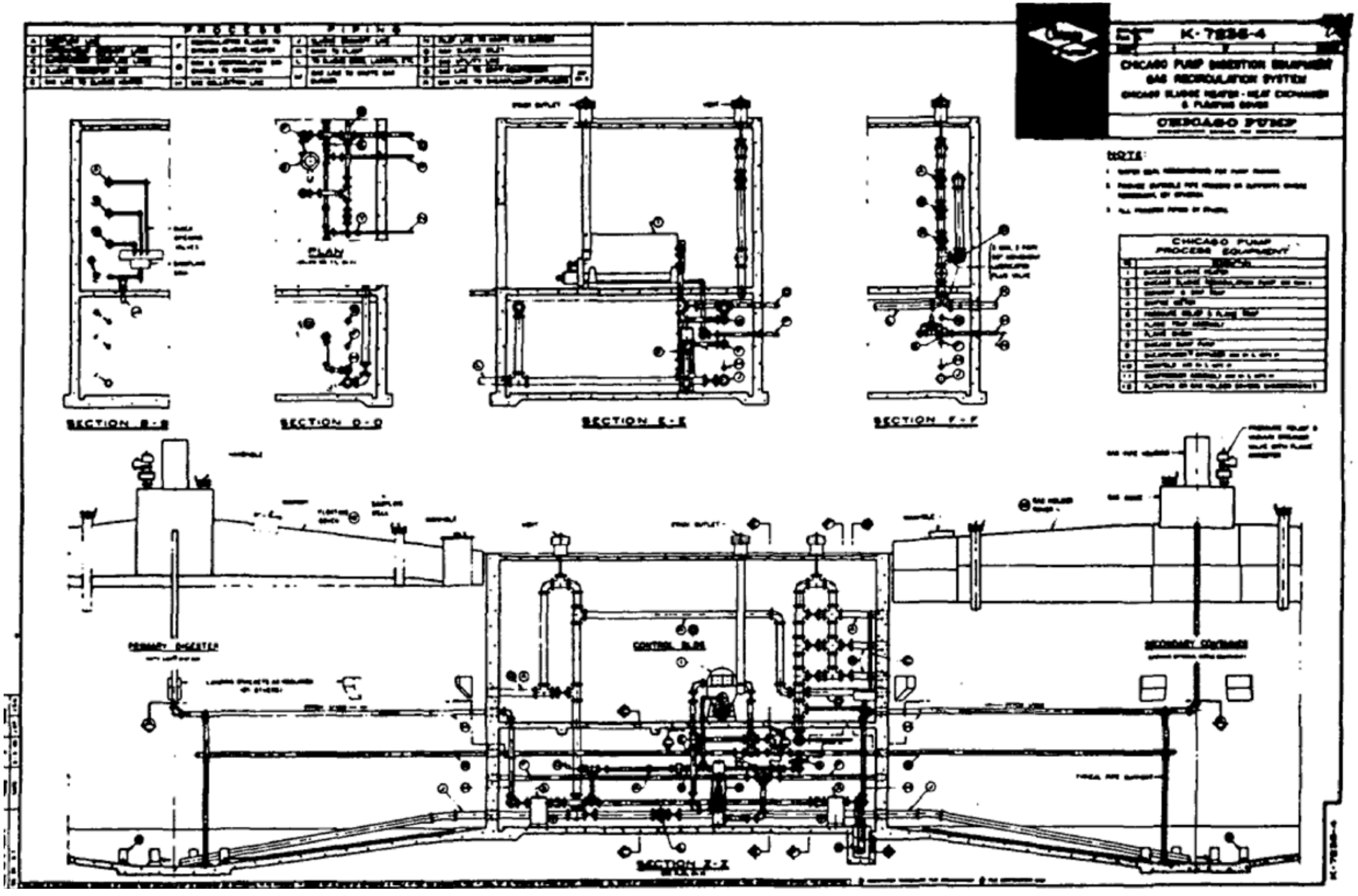


Fig. 15. Gas recirculation system.

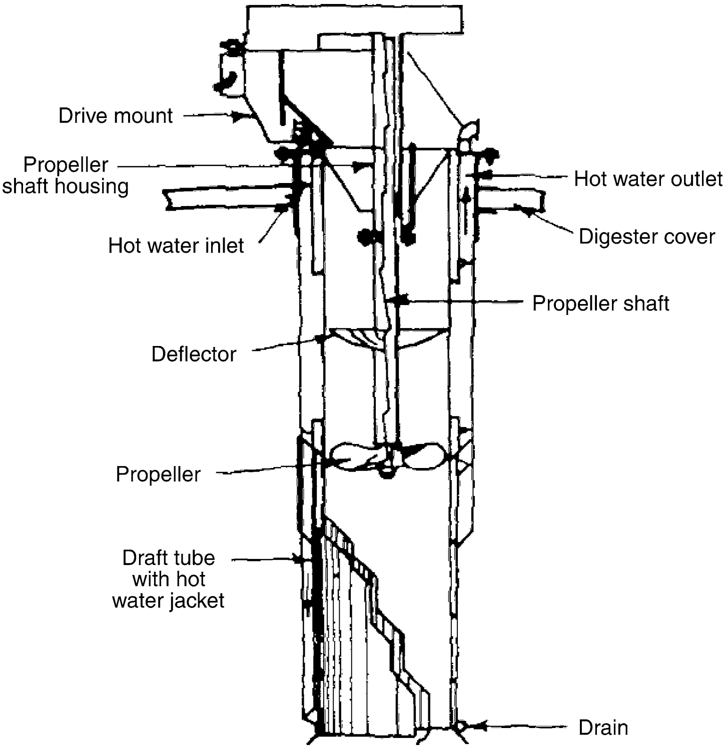


Fig. 16. Draft tube-type mixer.

3.4.3. Heating Systems

In most anaerobic digestion systems, it is necessary to supply heat from an external source in order to reach the desired operation temperature of approx 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 the heat lost to the surrounding medium. Figure 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 the digesters is dependent upon the shape of the reactor and the type of construction used. Cylindrical digesters that have a diameter, equal to the depth are most efficient in terms of heat retention. Different materials of construction have different thermal transfer coefficients. Heat losses from the digesters can be approximated from the following equation:

$$Q = UA(T_2 - T_1) \tag{13}$$

where Q is the heat loss from the tank (w [Btu/h]); U is the heat transfer coefficient (w/m^2/°C [Btu/h/ft^2/°F]); A is the surface area of tank element (m^2 [ft^2]); T_1 is the temperature outside the tank (°C [°F]); and T_2 is the 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

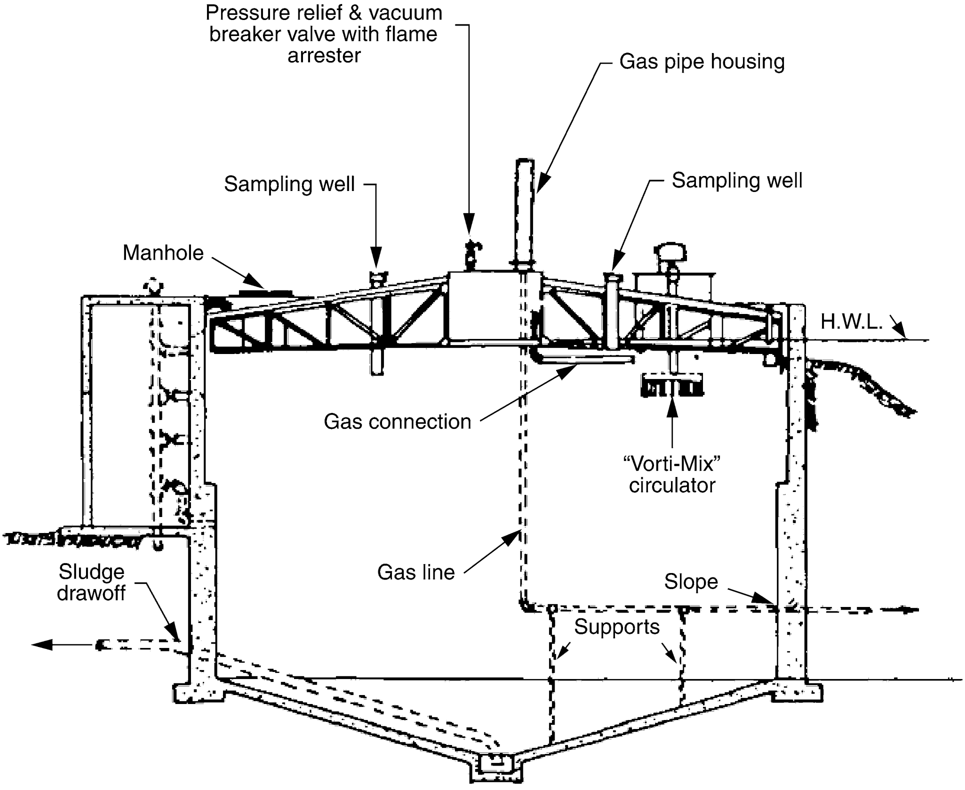


Fig. 17. Turbine-type mixing system.

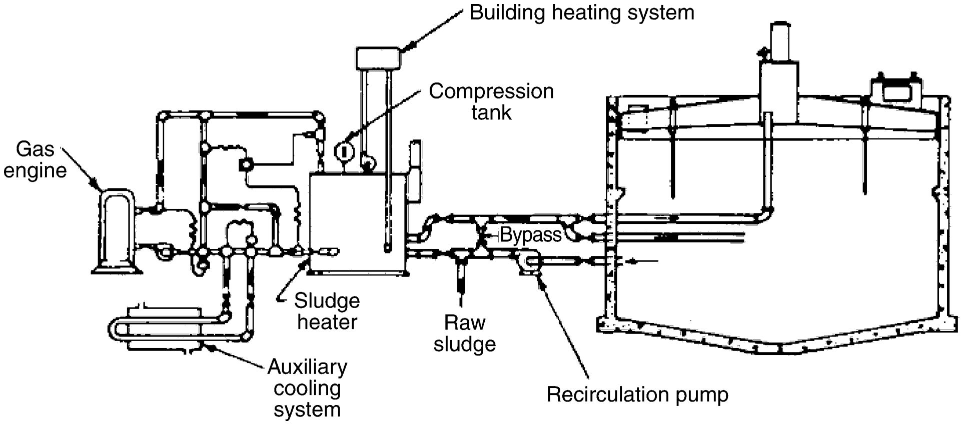


Fig. 18. Sludge heating system schematic.

coefficients for different digester sections presented in Table 4 can be used to calculate heat losses with Eq. (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-wk period expected should be used for the temperature outside the tank (T_1).

Table 4
Digester Heat Transfer Coefficients[a]

Digester section	W/m²/°C	Btu/h/ft²/°F
150 mm (6 in.) concrete roof	2.84	0.50
Floating cover with buildup 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] From ref. 16.

Table 5
Estimated Heat Losses From Anaerobic Digesters

	Heat losses					
	W/100 m³			Btu/h/1000 ft³		
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

An easier method of estimating the digester heat losses that does not require the consideration of heat losses through each element of the digester has been used. Approximately 2720 W/l00 m³ (2600 Btu/1000 ft³/h) will be lost from a well-insulated unmixed digester in the northernmost part of the United States. The values in Table 5 can be used to estimate the 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 is the amount of heat required (J [Btu]); C is the mean specific heat of raw sludge = 4200 J/kg/°C (1 Btu/lb/°F); w is the weight of sludge entering the tank per hour (kg [lb]); T_2 is the temperature of sludge in the tank (°C [°F]); and T_1 is the temperature of raw sludge entering tank (°C [°F]).

The average temperature of the raw wastewater during the coldest 2-wk 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

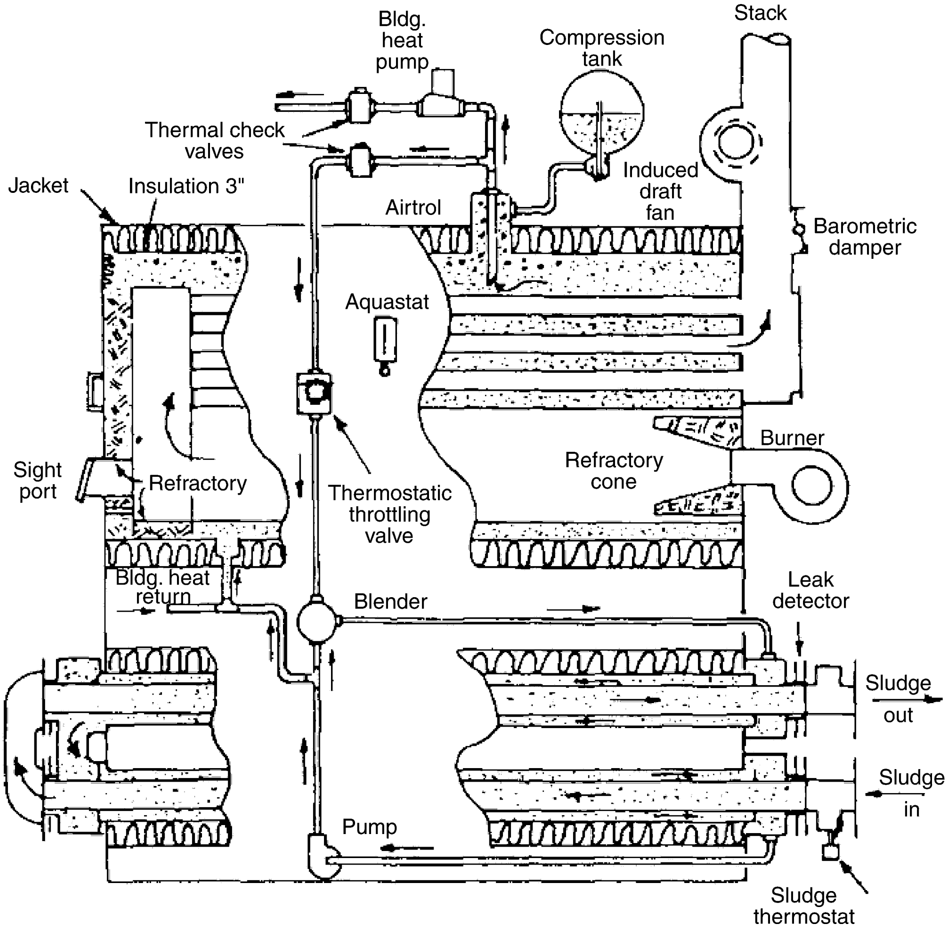


Fig. 19. External heat exchanger for use in anaerobic digesters.

coils. The latter method is generally considered to be outdated technology and the method of choice normally would be the use of external heat exchangers. Figure 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 the 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 maintain the desired level of mixing.

An example of a jacketed draft tube mixer is illustrated in Fig. 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 approx 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 the 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 earlier 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 can then be utilized 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.

Some gas storage is normally provided under the digester cover, as discussed previously. Storage is necessary to balance the demand for gas used to fuel utilization equipment with gas production from the digesters. In addition to the gravity-type gas holders similar to the floating digester covers, pressure-type holders are also used. Operating pressures used range from 1.4 to 7.0 kg/cm^2 (20–100 psi). Gas is pumped into 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 approx 3.5 m/s (11.5 ft/s) to avoid high-pressure losses and carryover of moisture from the condensate traps. Gas piping should be sloped to 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 the proper operation.

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 the gas lines. Flame traps should be installed in all gas lines that connect to the gas utilization 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 the particulates carried over in the gas from the digester, scale 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 utilization 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. As the 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 utilization equipment. Design pressure in the gas system and digester cover normally is approx 150–250 mm of water column (6–10 in.). Figure 20 shows a schematic of a complete digester gas system.

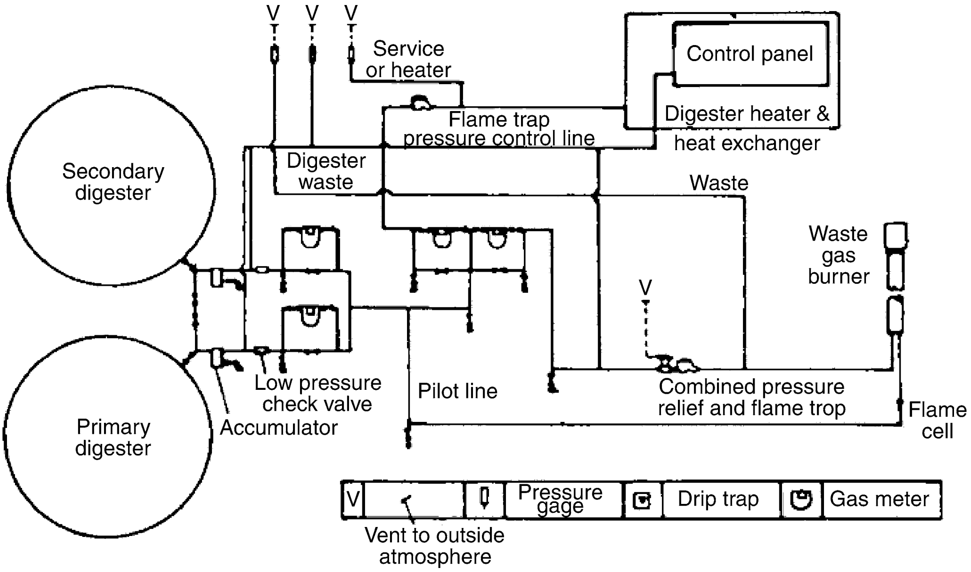


Fig. 20. Gas piping schematic of a modern anaerobic digestion system.

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 the digester gas may accumulate. Electrical fixtures in these areas should comply with the National Fire Protection Association requirements for hazardous locations.

3.5. Gas Utilization

Utilization 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 approx 5850 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. Utilization 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 utilize 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 the 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 H_2S. Removal of carbon dioxide will increase the heat value of the fuel, and equipment to accomplish this task may be justified in some instances. H_2S is a particular problem as it forms a corrosive liquid vapor when burned in combination with water vapor. The maintenance of boiler water temperatures more than 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.

Hot water boilers and internal combustion gas engines generally require a gas pressure of 76–130 mm (3–5 in.) of water for proper operation and to ensure a positive pressure throughout the gas system. Gas turbine engines require a fuel pressure of 10.5–14.1 kg/cm^2 (150–200 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 transferred 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, and so on), whereas positive displacement pumps (progressing cavity, peristaltic, and so on) 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.

Heat losses resulting from pipe friction can be significantly higher than those expected for water, so the designer must adjust head loss calculations accordingly. Brisbin (14) and Chou (15) investigated the flow of wastewater sludge in pipes and found that the Hazen-Williams C value varies as a function of the 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 (16,17).

Piping flexibility is extremely important, in order 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.

VS 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 TS 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 (<5000–7500 TS mg/L). The use of chemicals, such as polymers, may be warranted in some cases in order to obtain satisfactory supernatant quality so as not to cause upsets in the plant. Preparation of supernatant is often 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 approx 20 kg of actively digesting VS in the digester for each 1 kg of raw VS 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 in the maintenance of a relatively constant temperature as long as it is in the normal range of 30–37.5°C (86–100°F). Temperature changes should be kept to be 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. VS loadings and VS reduction in the digester.

All of the aforementioned parameters 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 the 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 (15) and the US EPA Operations Manual on Anaerobic Sludge Digestion (18). The material contained in these two publications is extremely helpful to the designers of unit processes, and it should be reviewed early in the design stage of a project in order to design for optimal operability of the facility.

5. CAPITAL AND OPERATING COSTS

5.1. Generals

The information on costs presented in this section must be used with care. Cost data which are included in Fig. 21 represents average costs for capital construction and operation and maintenance (19). 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.

a. **Capital costs.** 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 utilization of this gas for power recovery.
b. **Operating and maintenance costs.** Labor represents the most significant operating and maintenance cost for anaerobic digestion. This process requires a high degree of operating

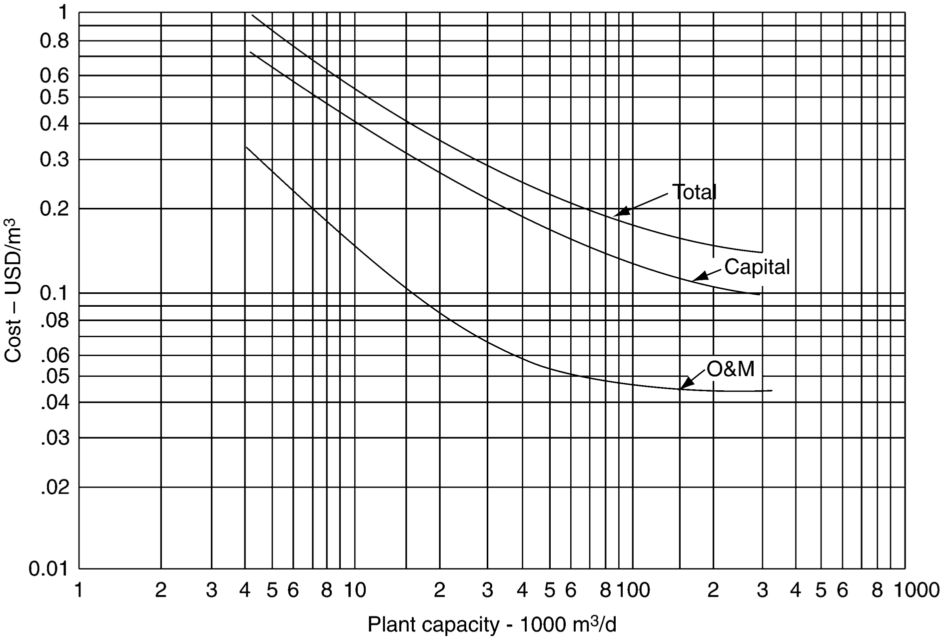


Fig. 21. Estimated costs of anaerobic digestion facilities.

control and supervision for peak efficiency. 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 operating and maintenance 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 0.1 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% TS (95% moisture) and that the waste secondary sludge will be thickened to 4% TS (96% moisture). The digested sludge should contain 8% TS (92% moisture). All sludges are assumed to have a specific gravity of 1.02 and the raw sludge contains 75% VS. Other pertinent design assumptions are as follows:

a. Only the primary digester will be mixed and heated to 35°C.
b. Sludge contains adequate nitrogen and phosphorus for biological growth.
c. The design loading is 1.28 kg/VS/m^3/d in accordance with GLUMRB standards.

Solution

a. Compute the weight of VS to be added to the digester daily.
Primary sludge (kg/d) = (40,000 persons) (0.10 kg/cap/d) (0.55)
= 2200 kg/d
= (2200 kg/d) (2.2046 lb/kg) = 4850 lb/d
Waste sludge (kg/d) = (40,000 persons) (0.05 kg/cap/d)
= 2000 kg/d (4409 lb/d)
VS to digester (kg/d) = (2200 + 2000) (0.75)
= 3150 kg VS/d (6945 lb VS/d).

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 m^3/d
= 43.1 m^3/d (264.172 gal/m^3) = 11,386 gpd

$$\text{Waste sludge } (\text{m}^3/\text{d}) = \frac{2200\,\text{kg/d}}{(0.04)(1.02)(1000\,\text{kg/m}^3)}$$

= 49 m^3/d (12,944 gpd)
Sludge to digester (m^3/d) = 43.1 m^3/d primary + 49 m^3/d waste
= 92.1 m^3/d (24,330 gpd).

c. Compute the volume of the first stage digester.

$$\text{Volume } \left(\text{cm}^3\right) = \frac{3150\,\text{kg VS/d}}{1.28\,\text{kg VS/m}^3 \cdot \text{d)}}$$

= 2461 m^3
= 2461 m^3 (35.3147 ft^3/m^3) = 86,909 ft^3
(The volume of the second stage digester also should be 2461 m^3)

d. Compute the hydraulic residence time in the primary digester using Eq. (9).

$$\theta = \theta_c = V/Q$$

$$\theta_c = \frac{2461\,\text{m}^3}{92.1\,\text{m}^3/\text{d}} = 26.7\text{ d} > 10\text{ d minimum}$$

e. Estimate the quantity of digested sludge produced.
From Fig. 7, θ = 27 d, VS = 75%, then VS reduction (VSR) = 54%.
Weight of total = weight of fixed solids (FS) + weight of volatile solids (VS)
= (2200 + 2000) (0.25) + (2200 + 2000) (0.75) (1 – 0.54)
= 1050 kg FS/d + 1449 kg VS/d
= 2499 kg/d (5509 lb/d) of digested sludge on a dry weight basis

$$\text{Volume of digested sludge} = \frac{2499\,\text{kg/d}}{(0.08)(1.02)(1000\,\text{kg/m}^3)}$$

$$= 30.6\ \text{m}^3\text{/d (8084 gpd)}$$

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 5:
Heat loss in digester $= (4190\ \text{W}/100\ \text{m}^3) \times 2461\,\text{m}^3$

$= 1.03 \times 10^5$ W
$= 1.03 \times 10^5$ W (0.0009483 Btu/s-W) (3600 s/h) $= 3.51 \times 10^5$ Btu/h

From Eq. (14):
$H = WC\,(T_2 - T_1)$

$$H = \left(\frac{92.1\ \text{m}^3/\text{d} \times 1.02 \times 1000\ \text{kg/m}^3}{24\ \text{h/d}}\right)(4200\ \text{J/kg/°C})\ (35\text{°C} - 10\text{°C})\left(\frac{1\ \text{W}}{3600\ \text{J}}\right)$$

$= 1.14 \times 10^5$ W (3.89×10^5 Btu/h)
Total heat requirement $= 1.03 \times 10^5 + 1.14 \times 10^5$
$= 2.17 \times 10^5$ W (7.41×10^5 Btu/h)

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-stages, digesters required to treat the sludge from a community of 80,000 persons. The wastewater flow is assumed to be 30,284 m^3/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% TS (94% moisture) and that the thickened waste secondary sludge contains 4% TS (96% moisture). The digested sludge should contain 8% TS (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 VS to be added to the digester daily.

$$\text{VS in primary sludge (kg/d)} = \frac{(275\ \text{mg/L} - 125\ \text{mg/L}) \times 30{,}284\ \text{m}^3/\text{d} \times 0.75\ (1000\ \text{L/m}^3)}{1{,}000{,}000\,\dfrac{\text{mg}}{\text{kg}}}$$

= 3407 kg VS/d (7511 lb VS/d)

$$\text{BOD}_5 \text{ to aerator (kg/d)} = \frac{165\ \text{mg/L} \times 30{,}284\ \text{m}^3/\text{d}}{1{,}000{,}000 \dfrac{\text{mg}}{\text{kg}}} \left(1000 \frac{\text{L}}{\text{m}^3}\right)$$

= 4997 kg/d (11,016 lb/d)

$$\text{VSS in waste sludge (kg/d)} = \left(\frac{4997\ \text{kg BOD}_5}{\text{d}}\right)\left(\frac{0.28\ \text{kg VSS}}{\text{kg BOD}_5}\right)$$

$$= \frac{1399\ \text{kg VSS}}{\text{d}}\ \ (3084\ \text{lb VSS/d})$$

VS to digester (kg/d) = 3407 + 1399
= 4806 kg VS/d (13,995 lb VS/d).

b. Compute the volume of raw sludge expected.

$$\text{Primary sludge (m}^3/\text{d)} = \frac{3407\ \text{kg VSS/d}}{(0.06)(0.75)(1.02)(1000\ \text{kg/m}^3)}$$

= 74.2 m^3/d (19,601 gpd)

Assume the waste sludge is 75% volatile.

$$\text{Waste sludge (m}^3/\text{d)} = \frac{1399\ \text{kg VSS/d}}{(0.04)(0.75)(1.02)(1000\ \text{kg/m}^3)}$$

= 45.7 m^3/d (12,073 gpd)

Sludge to digester (m^3/d) = 74.2 m^3/d primary + 45.7 m^3/d waste
= 119.9 m^3/d (31,673 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% TS.

c. Compute the volume of the first stage digester.
From Table 3, T = 35°C, θ_c = 10 d for design
From Table 1, θ_c = 10 d, TS = 4.47%, by interpolation
The VS loading factor is 3.42 kg/m^3/d

$$\text{Volume in m}^3 = \frac{4806\ \text{kg VS/d}}{3.42\ \text{kg/m}^3/\text{d}}$$

= 1405 m^3 (49,616 ft^3)

(The volume of the second stage digester also should be 1405 m^3).

d. Check the hydraulic residence time using Eq. (9)

$$\theta = \theta_c = V/Q$$

$$\theta_c = \frac{1405\ \text{m}^3}{119.9\ \text{m}^3/\text{d}} = 11.7 \text{ OR } 12\ \text{d}$$

e. Estimate the quantity of digested sludge produced.
From Fig. 8, $\theta = 12$ d, VS = 75%, then VS reduction (VSR) = 57%
Weight of total = weight of fixed + weight of VS

$$\text{Weight of TS in combined raw sludge} = \frac{4806\ \text{kg VS/d}}{0.75}$$

$$= 6408\ \text{kg/d}\ (14{,}127\ \text{lb/d})$$

Weight of TS in digested sludge = (6408) (0.25) + (6408) (0.75) (1 − 0.55)
= 1602 kg FS/d + 2163 kg VS/d
= 3765 kg/d (8300 lb/d) of digested sludge on a dry weight basis.

$$\text{Volume of digested sludge} = \frac{3765\ \text{kg/d}}{(0.08)\ (1.02)\ (1000\ \text{kg/m}^3)} = 46\ \text{m}^3\text{/d}\ (12{,}152\ \text{gpd}).$$

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) and that the gas has an energy value of 5850 kg-cal/m^3
= (5850 kg-cal/m^3) (3.968 Btu/kg-cal) (35.31 m^3/ft^3)
= 657 Btu/ft^3
VS destroyed (kg/d) = (6408 kg TS/d)(0.75)(0.55)
= 1202 kg VS/d
Gas produced (m^3/d) = (1202 kg VS/d) (0.90 m^3/kg VS)
= 1082 m^3/d (38,210 ft^3/d)
Energy content of gas produced = (1082 m^3/d) (5850 kg-cal/m^3)
= 6.33×10^6 kg-cal/d
= (6.33×10^6 kg-cal/d) (3.968 Btu/kg-cal)
= 2.51×10^7 Btu/d.

The digester heating requirements would be estimated in the same manner as shown in Section 6.1. 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 = 1500 m^3/d (396,300 gpd)
BOD_5 concentration = 1200 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. $\theta_c = 14$ d (see Table 3).
c. Efficiency of waste utilization $E_s = 0.80$.
d. The waste contains adequate nitrogen and phosphorus for biological growth.
e. $Y = 0.07$ kg VS/kg BOD_5 utilized and $k_d = 0.03$/d at 30°C.

Solution

1. Compute the daily BOD_5 loading.

$$\text{BOD}_5 = \left(1200\ \frac{\text{mg}}{\text{L}}\right)\left(\frac{\text{kg}}{10^6\ \text{mg}}\right)\left(\frac{1500\ \text{m}^3}{\text{d}}\right)\left(\frac{1000\ \text{L}}{\text{m}^3}\right)$$

$$= 1800\ \text{kg BOD}_5\text{/d}\ (3968\ \text{lb/d})$$

2. Compute the digester volume using Eq. (9)

$$\theta = \theta_c = V/Q$$

$$\theta_c = 14 \text{ d}$$

$$V = (1500 \text{ m}^3/\text{d})(14 \text{ d})$$

$$= 21{,}000 \text{ m}^3 \ (741{,}594 \text{ ft}^3)$$

Several digesters in parallel will be needed meet this volume requirement.

3. Computing the volumetric loading:

$$\text{BOD}_5 \text{ kg/m}^3\text{/d} = 1800 \text{ kg/d}/21{,}000 \text{ m}^3 = 0.09 \text{ kg/m}^3\text{/d}$$

4. Compute the effluent waste concentration using Eq. (4).

$$E_s = \frac{100(S_\text{o} - S_1)}{S_\text{o}}$$

$$S_1 = \frac{100(1200) - 80(1200)}{100}$$

$$= 240 \text{ mg/L}$$

5. Computing the quantity of VS produced per day using Eq. (8).

$$X = \frac{Y\ Q\ (S_\text{o} - S_1)}{1 + (k_d)(\theta_c)} \tag{8}$$

$$= \frac{\left(0.07 \dfrac{\text{kg VS}}{\text{kg BOD}_5}\right)\left(1500 \dfrac{\text{m}^3}{\text{d}}\right)\left(1000 \dfrac{\text{L}}{\text{m}^3}\right)\left(1200 \dfrac{\text{mg}}{\text{L}} - 240 \dfrac{\text{mg}}{\text{L}}\right)\left(\dfrac{\text{g}}{1000 \text{ mg}}\right)\left(\dfrac{\text{kg}}{1000 \text{ g}}\right)}{1 + (0.03 \text{ d}^{-1})(14 \text{ d})}$$

$$= 71 \text{ kg/d } (156 \text{ lb/d})$$

Using Eq. (3) to determine the volume (m^3) of methane gas produced per day and letting X in Eq. (8) equal to A.

$$C = 0.35 \text{ (eF-1.42A)}$$

$$= 0.35([0.80]\ [1800] - 1.42\ [71])$$

$$=0.35(1440 - 100.82)$$

$$= 469 \text{ m}^3\text{/d } (16{,}563 \text{ ft}^3\text{/d})$$

6. 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 operate as both suspended growth and attached growth reactor. The AF utilizes 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% (20–23). 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 (22,24). Recycling of the effluent is typically utilized to maintain a uniform hydraulic loading to the treatment unit. The flow through for an AF can either be up-flow or down-flow. AF operated in the up-flow pattern, generally has more biomass in suspension than AF operated in the down-flow pattern. With either flow pattern, gas is collected in the top of the reactor.

During startup, Punal et al. (20) determined that limiting nitrogen concentration in the influent during the first 2 wk followed by a nitrogen balance influent promotes bacterial biofilm formation. Snow and Tay (21) 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 approx 75% COD removal. During startup of AF, Smith et al. (22) showed that the up-flow velocity should be in the range of 8–10 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 up-flow 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 has 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 with media with larger void volume. This is illustrated by the specific surface area for the media ranging from 4000 to 10,000 m^2/m^3, but having void volumes ranging from 45 to 55% (23). Silicon sand or granular activated carbon are commonly used as media (23,25). The flow pattern for anaerobic EB reactor is upwards at velocities that are sufficient to expand the bed of media by 10–30%, whereas in a FB reactor the up-flow velocity is increased to expand the media from 25 to 300%. In the EB reactor, the media with biofilm 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 biofilm 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, solid 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. (25) operated a commercial-scale AFB reactor filled with granular activated carbon (GAC). The unit was 0.5 m in diameter and 4.6 m tall. The AFB was operated at steady-state flows of 0.03–0.05 l/s, which provide averaged hydraulic retention time of 10 h. 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, 2410 mg/L of alcohol and 9200 mg/L of COD. The results of this study showed that the AFB with GAC can lower the concentration of difficult to degrade organic DNT to less than 0.08 mg/L or greater than 99% removal.

A suspended growth process that has been developed is the up-flow anaerobic sludge blanket (UASB; 10,23,24,26–28). 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, density 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 of wastewater being treated can cause the makeup of the blanket to vary. Wastewater containing suspended solids or other matter that are not being trapped by the granular biomass will pass through and form a less dense blanket above denser granular biomass blanket or to 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 gases separator which allows solids to be returned to the blanket and solids/gases 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 results in the increase in the depth of the blanket. As depth of the blanket increases because of solids buildup, the effectiveness of the separator will decrease. This buildup can interfere with solids and gases 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 (28) reported that optimum granulation occurred when the calcium concentration ranged from 150 to 300 mg/L when COD influent concentration of 4000 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, because 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 d and the range reached 1–2 mm after 60 d 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

Table 6
Performance Data for Anaerobic Processes

Process	Hydraulic detention time (h)	Organic loading (kg COD/m^3/d)	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

calcium in 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 up-flow anaerobic sludge blanket process (23,29).

Zoutberg and de Been (29) reports on the full-scale EGSB installation treating formaldehyde operated at superficial up-flow velocity in the reactor at 9.4 m/h, which is much high than conventional UASB reactor maximum superficial up-flow velocity of 1 m/h. This high 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 approx 5000 to 45,000 mg/L and effluent ranged from 350 to 900 mg/L. This EGSB achieved 98% removal of COD. Typical performance data for anaerobic processes are presented in the Table 6. These values are approximate and are intended to give a range that is applicable for the process.

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 (30–32,44).

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 ESD is an example to this type of digester. In mid-1970s, the first ESD were built in the United States and are 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 37–45°conical side slopes (32). Both designs provide small bottom area and the steep 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 (31,32). 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; 33–35,36,37,46,47) has been developed to meet the requirements for pathogen reduction (density of fecal coli form in the bio solids must be less than 1000 most probable number per gram of TS) and vector attraction reduction (38% reduction in VS content of the bio solids) as required in Title 40 of the Code of Federal Regulations (CFR), Part 503 (38). 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 (SRT's) for the stages, a balance must be obtained between the pathogen reduction, heat exchanger size, and energy consumption. As reported by Han and Dague (36), 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 coliform, when the SRT varied from 3.3 to 5 d with a constant operating temperature of 55°C. Varying SRT for the mesophilic phase (35°C) from 6.7 to 10 d, the researchers observed that TPAD process achieved 39% VS destruction at SRT of 6.7 d and 53.2% VS destruction at SRT of 10 d. The research also reported that a single stage mesophilic process (35°C) with SRT of 10 d achieved 32% volatile destruction and with SRT of 15 d 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 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 (39), and developed control system to regulate the COD (40), respectively. Mathematical model has also been developed for predicting the conversion of complex organic into biogas in a batch reactor (41).

Taricska (39) 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 approx 900, 2300, and 3800 mg/L. The anaerobic unit with media had hydraulic detention time of 6.5 d and the aerobic units with media had hydraulic detention times of 5, 10, and 15 d. 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.

Table 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 and Farrell (35)]	WAS	7	27	43	
Wilhelmshaven, Germany [Schafer and Farrell (35)]	Primary and WAS	3–5	13–15	54	
Belmont WWTP, Indiana [Shimp, et al. (33)]	Primary and WAS	2	10	>38	<1000
[Chao et al. (36)]	Pulp and Paper	2.5–5	7.5–15	38.2	
[Vanderburgh and Ellis (46)]	Primary and WAS			>38	<1000
[Oles et al. (47)]	Municipal	2–3	12–15		

Mendez-Acosta et al. (40) proposed a control system for anaerobic digestion. The proposed system is a dynamic output feedback control for the regulation of the 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.

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. The readers are referred to the literature (27,33,45–49) for the latest developments in anaerobic digestion.

NOMENCLATURE

A	Kg VS produced/d
A	Surface area of tank element (m^2 [ft^2])
A	Excess microorganism production rate (g/d)
C	Mean specific heat of raw sludge (4200 J/kg/°C [1 Btu/lb/°F])
C	Methane produced (m^3/d)
e	Efficiency of waste utilization
E_S	Process efficiency (%)
F	BOD$_5$ added (kg/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 utilization per unit weight of microorganism is one half the maximum rate (g/L)
Q	Heat loss from the tank (W [Btu/h])
Q	Waste flow rate (m^3/d)
Q_r	Return sludge flow rate (m^3/d)
r	Return sludge ratio (Q_r/Q)
S_0	Influent substrate concentration (g/L)
S_1	Effluent substrate concentration (g/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	Food to microorganism ratio
U	Heat transfer coefficient ($W/m^2/°C$ [$Btu/h/ft^2/°F$])
V	Volume of the reactor (m^3)
w	Weight of sludge entering the tank per hour (kg [lb])
X	Mass of volatile solids in reactor (g)
X_r	Mass of volatile solids in the return sludge (g)
Y	Cell yield (g/g)
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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6
Aerobic Digestion

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CONTENTS

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.

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Studies on aerobic digestion of municipal wastewater sludge have been conducted since the early 1950s (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 approx 130 installations in plants with flow from 10,000 to 100,000 gpd (37.8–378 m^3/d). By the late 1960s and early 1970s, consulting engineers across the country were specifying aerobic digestion facilities for many of the plants they were designing. As of early 1980s, 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 cellular 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):

$$\text{Organic matter} + O_2 \rightarrow \text{Cellular material} + CO_2 + H_2O \qquad (1)$$

$$\text{Cellular material} + O_2 \rightarrow \text{Digested sludge} + CO_2 + H_2O \qquad (2)$$

The process described by Eq. (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):

a. Have capital costs generally lower than for anaerobic systems for plants less than 5 MGD (220 L/s).
b. Are relatively easy to operate compared with anaerobic systems.
c. VSS is reduced to 40–50%, nearly equivalent to that for anaerobic.
d. Do not generate nuisance odor.
e. Will produce a supernatant low in BOD_5, suspended solids, and ammonia nitrogen.
f. Reduce the quantity of grease in the sludge mass.
g. A relatively stable humus like end product is produced.
h. Reduce the number of pathogens to a low level under normal design. Under autoheated thermophilic design, many systems provide 100% pathogen destruction.

2.3. Disadvantages

As with any process, there are also certain disadvantages. In aerobic digestion processes, the disadvantages are as follows:

a. Usually produce a digested sludge with very poor mechanical dewatering characteristics.
b. Have high power costs to supply oxygen, even for very small plants.

c. Are significantly influenced in performance by temperature, location, and type of tank material.
d. No heavy metal removal.
e. Lack of useful byproduct (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 Semibatch Operation

Originally, aerobic digestion was designed as a semibatch 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 2–3 wk 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 2 and 4%. 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 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 semibatch 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 Fig. 1. As in the semibatch 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 metal 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 1–2 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 autoheated 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 (>45°C) using air as the source of oxygen to aerate the sludge. The operation is

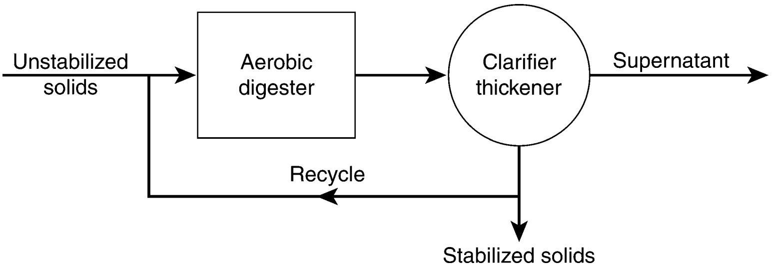

Fig. 1. Flow diagram for a conventional aerobic digestion process (*Source:* US EPA) (9).

autothermal, i.e., 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 more than 4%. 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 d 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 45–70°C and the recommended retention time is from 2 to 10 d (30).

Advantages claimed for this mode of operation are (13,21,30–32) as follows:

a. Higher rates of organic solids destruction.
b. Smaller volume requirements.
c. Production of a pasteurized sludge.
d. Destruction of all weed seeds.
e. 30–40% less oxygen requirement.
f. Improved solids–liquid separation through decreased liquid viscosity.

Disadvantages cited for this process are that:

a. It must incorporate a thickening operation.
b. Mixing requirements are higher because of the higher solids content.
c. Nonoxygen-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 refs. 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 (>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 the 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 >45°C a high degree of digestion can be obtained with less than 5 d 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 (WAS) because of the large amount of heat required to raise WAS at 0.5% solids to thermophilic temperatures. The process has high operating cost (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 5 d or less and at operating temperatures in the thermophilic range of 45–60°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

As 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 lower rather than on a high grade, and using subsurface

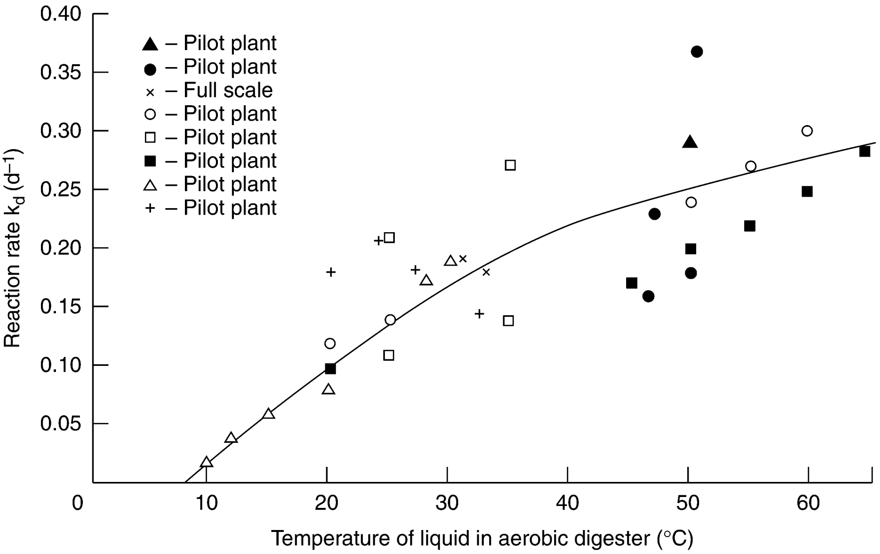

Fig. 2. Aerobic digestion reaction rate (k_d) as a function of sludge temperature (*Source:* US EPA) (9).

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, although some studies (51,52) have shown that there may be destruction of the nonorganics 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:

$$\frac{dM}{dt} = -K_d\, M \tag{3}$$

where dM/dt is the rate of change of biodegradable volatile solids per unit of time (mg/L/d); K_d is the reaction rate constant (d^{-1}); M is the concentration of biodegradable volatile solids remaining at time t in the aerobic digester [mg/L (ppm)]; t is the time (d).

The time t in Eq. (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 more 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 pseudoconstant, because the term's value is the average result of many influences (9). Figure 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.

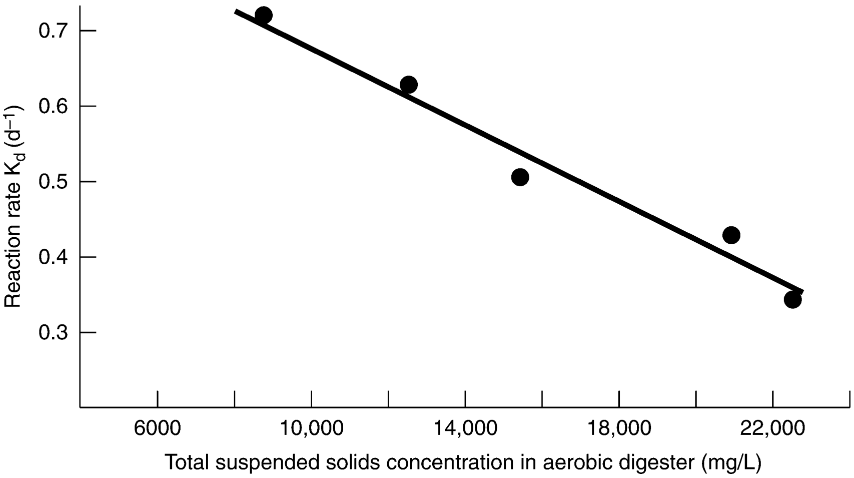

Fig. 3. Aerobic digestion reaction rate (k_d) as a function of solids concentration (*Source:* US EPA) (9).

Little research has been conducted on the effect of solids concentration on reaction rate K_d. The results of one study (49) with WAS at a temperature of 68°F (20°C) are shown on Fig. 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 15–20 d). Up to 80% of the cell tissue may be oxidized; the remaining fractions contain inert and nonbiodegradable materials (9).

4.3. Oxygen Requirements

Activated sludge biomass is most often represented by the empirical equation $C_5H_7O_2N$. The aerobic digestion of the sludge is usually depicted as follows:

$$C_5H_7O_2N + 5O_2 = 5CO_2 + 2H_2O + NH_3 \tag{4a}$$

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 Eq. (4b):

$$C_5H_7O_2N + 7O_2 \rightarrow 5CO_2 + 3H_2O + H^+ NO_3^- \tag{4b}$$

Hypothetically, this equation indicates that 1.98 lb (0.898 kg) of oxygen are required to oxidize 1 lb (0.45 kg) of cell mass. However, from pilot and full-scale studies, the pounds of oxygen required to degrade a pound of volatile solids were found to be 1.74–2.07 (0.789–0.939 kg). For mesophilic systems, a design value of 2 is recommended. For autothermal systems, which have temperatures higher than 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, lb oxygen per 1000 lb 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 in Fig. 4. Specific oxygen utilization is seen to decrease with increase in sludge age and decrease in digestion temperature (9).

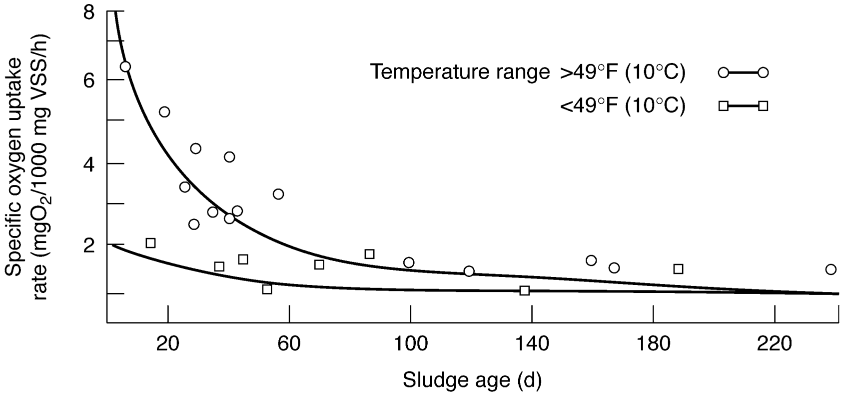

Fig. 4. Effect of sludge age and temperature on oxygen uptake rates (*Source:* US EPA) (9).

Field studies have also indicated that a minimum value of 1 mg of oxygen per liter should be maintained in the digester at all times (14).

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 hp/1000 ft^3 of tank volume (13–106 kW/1000 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 in Fig. 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 aerobically digested sludge. As an example, a designer

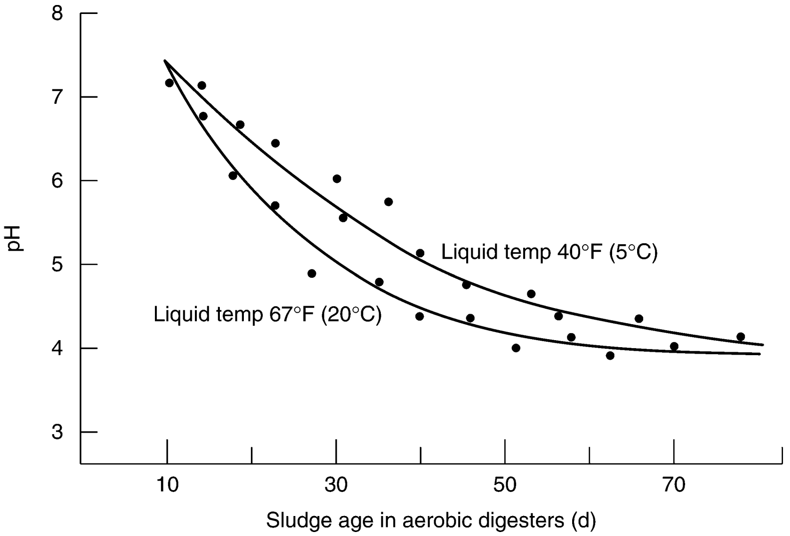

Fig. 5. Influence of sludge age on pH during aerobic digestion (*Source:* US EPA) (9).

would probably consider designing a rotary vacuum filter for a production rate of 1.5 lb of dry solids/ft^2/h (7.4 kg/m^2/h), a cake solids concentration of 16%, 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% 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 6 is a plot of volatile solids reduction against the parameter degree-days. Data were taken from both pilot and full-scale studies on several types of municipal wastewater sludges (9). Figure 6 indicates that, for these sludges, volatile solids reductions of 40–50% are obtainable under normal aeration conditions.

Up to 85% 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 100–12,000 (mg/L).
b. BOD_5 50–1700 (mg/L).
c. Soluble BOD_5 4–200 (mg/L).
d. COD 200–8000 (mg/L).
e. Kjeldahl N (TKN) 10–400 (mg/L).
f. Total P 20–250 (mg/L).
g. Soluble P 2–60 (mg/L).
h. pH 5.5–7.7.

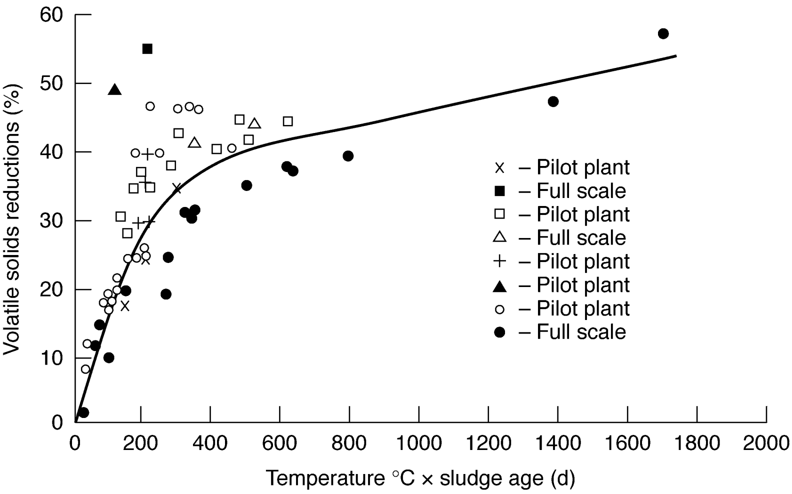

Fig. 6. Volatile solids reduction as a function of sludge temperature and age (*Source:* US EPA) (9).

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 refs. 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: 18–20 d at 20°C for mixed sludges from activated sludge or trickling filter plant, 10–16 d for WAS only, 16–18 d average for activated sludge from plants without primary settling; volume allowance, 3–4 ft^3/capita; VSS loading, 0.02–0.4 lb/ft^3/d; air requirements, 20–60 ft^3/min/1000 ft^3; minimum DO, 1–2 mg/L; energy for mechanical mixing, 0.75–1.25 hp/1000 ft^3; oxygen requirements, 2 lb/lb of cell tissue destroyed (includes nitrification demand), 1.6–1.9 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 (%).
d. Secondary solids content (%).
e. Specific gravity.
f. Volatile solids content (%).

6.2. Design Parameters

The current design criteria for aerobic digesters are summarized in Tables 1 and 2.

6.3. Design Procedure

6.3.1. Calculate Total Quantity of Raw Sludge

$$Q = W_s\,(100)/(\text{specific gravity})\ (\%\ \text{solids})\ (8.34) \qquad (5)$$

Table 1
Aerobic Digester Design Criteria

Parameter	Value
Hydraulic detention time (days at 20°C[a])	
Activated sludge only	12–16
Activated sludge from plant operated without primary settling	16–18
Primary plus activated or trickling-filter sludge	18–22
Solids loading (lb volatile solids/ft^3/d)	0.1–0.2
Oxygen requirements	
BOD_5 in primary sludge (lb/lb cell tissue)	.2
Energy requirements for mixing	
Mechanical aerators (hp/1000 ft^3)	0.5–1
Air mixing (cfm/1000 ft^3)	20–30
Dissolved oxygen level in liquid (mg/L)	1–2

[a] Detention times should be increased for temperatures less than 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.

Table 2
US EPA Aerobic Digester Design Criteria

	Days	Liquid temperature (°F)
Solids residence time required to achieve:		
40% Volatile solids reduction	108	40
	31	60
	18	80
55% Volatile solids reduction	386	40
	109	60
	64	80
Oxygen requirements	2 lb of oxygen per lb of volatile solids destroyed when liquid temperature 113°F or less 1.45 lb of oxygen per lb of volatile solids destroyed when liquid temperature more than 113°F	
Oxygen residual	1 mg/L of oxygen at worst design conditions	
Expected maximum solids concentration achievable with decanting	2.5–3.5% 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 hp/1000 ft^3 of tank volume	

Source: US EPA (9)
1 lb = 0.454 kg.
1 hp/1000 ft^3 = 26.6 kW/1000 m^3.

where Q is the quantity of raw sludge production (gpd); W_s is the 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 is the volume of digester (gal); t is the hydraulic detention time (d); Q is the quantity of raw sludge production (gpd).

6.3.3. Check Volatile Solids Loading

$$L_{vs} = \frac{W_{vs}\ (7.48)}{V} < 0.1 - 0.2 \tag{7}$$

where L_{vs} is the volatile solids loading (lb/ft^3/d); W_{vs} is the weight of volatile solids production (lb/d); V is the volume of digester (gal).

6.3.4. Calculate Solids Retention Time

a. Assume percent destruction of volatile solids: 40% is common but it increases with temperature and retention time from approx 33 to 70%.

b. Calculate solids accumulation per day.

$$W_{ac} = W_s - W_s \text{ (volatile \%/100) (destroyed \%/100) (0.75)} \tag{8}$$

where W_{ac} is the solids accumulation (lb/d); W_s is the weight of solids in produced sludge (lb/d).

c. Assume MISS in digester and calculate total digester capacity.

$$W_{dc} = (V)(\text{MLSS})(8.34)(10^{-6}) \tag{9}$$

where W_{dc} is the digester capacity (lb); MLSS is the mixed liquor SS in digester (mg/L); V is the volume of digester (gal).

d. Calculate solids retention time.

$$SRT = \frac{W_{dc}}{W_{ac}} \tag{10}$$

where SRT is the solids retention time (d); W_{dc} is the digester capacity (lb); W_{ac} is the solids accumulation (lb/d).

6.3.5. Calculate Sludge Wasting Schedule

Assume solids content in digested sludge is approx 2.5%.

$$V_w = \frac{\text{Total sludge in digester (lb) (100)}}{\text{(specific gravity)(solids \%)(8.34)}} \tag{11}$$

where V_w is the 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.

$$O_2 = (O_r)\ W_s \text{ (volatile \%/100) (destroyed \%/100)} \tag{12}$$

where O_2 is the total oxygen required (lb/d); O_r is the oxygen required/lb of volatile solids destroyed = 2 lb; W_s is the weight of solids in produced sludge (lb/d).

6.3.7. Calculate Air Requirement for Mixing

a. Assume standard transfer efficiency (%).
b. Assume constants α, β, and ρ.
c. Select summer temperature (T).
d. Calculate operating transfer efficiency.

$$\varepsilon_o = \varepsilon_s\ [(C_s)_{\mathrm{T}}\ \beta\rho - C_L]\ \alpha\ 1.024^{T-20}/9.20 \tag{13}$$

where ε_o is the operating transfer efficiency (%); ε_s is the standard transfer efficiency (%) (5–8%); $(C_s)_T$ is the oxygen saturation at the summer temperature; β is the (C_s waste/C_s water) = 0.9; ρ is the correction for altitude = 1; C_L is the minimum oxygen to be maintained in the digester (mg/L); α is the (K_{La} waste/K_{La} water) = 0.9; K_{La} is the oxygen transfer coefficient; T is the temperature (°C).

e. Calculate air supply; check against a minimum of 20 cfm/1000 ft^3.

$$Q_{\mathrm{air}} = \mathrm{O}_2\ (7.48)\ (10^5)/(\varepsilon_o\%)\ (0.0176\ \mathrm{lb\ O_2\ /ft^3\ air})\ 1440\ (\mathrm{min/d})\ V \tag{14}$$

where Q_{air} is the air supply (cfm/1000 ft^3); O_2 is the total oxygen required (lb/d); ε_o is the operating transfer efficiency (%);V is the volume of the basin (gal).

6.4. Output Data

a. Raw sludge specific gravity.
b. Detention time (d).
c. Volatile solids destroyed (%).
d. Mixed liquor solids (mg/L).
e. Solids in digested sludge (%).
f. Rate constant, BOD_5 applied.
g. Coefficient of oxygen saturation in waste/oxygen saturation in water.
h. Standard transfer efficiency (%).
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/l000 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 refs. 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 2006 using US Army Corps of Engineers Civil Works Construction Cost Index System for utilities, Manual no. 1110-2-1304 (72, Appendix A), and is given in Eq. (15).

$$C = 3.53 \times 10^5\ Q^{1.14} \tag{15}$$

where C is the capital cost of process in 2006 USD; Q is the plant design flow (MGD).

Table 3
Aerobic Digestion Labor Requirement

	Labor man hours per year		
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	1500	300	1800

Source: US EPA (73)
1 MGD = 3786 m^3/d.

The associated costs included those for excavation, process piping, equipment, concrete, and steel. In addition, such costs as those for administration and engineering are equal to 0.2264 of the capital cost (9):

$$C_{A\&E} = 0.2264\ C \tag{16}$$

where $C_{A\&E}$ is the cost of administration and engineering in 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 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% 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 20–30% 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 5–40 gal/unit/change (19–152 L/unit/change). Furthermore, the designer must make sure an adequate inventory of spare parts is available.

8. RECENT DEVELOPMENTS AND SUMMARY

8.1. Recent Developments

Since 1990 there have been many new developments in both aerobic and anaerobic digestion processes (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.
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 2.5–10 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. The principal difference between VERTAD and conventional ATAD-Air and ATAD-Oxygen systems is that its in-ground 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 (>40%). This high rate of oxygen transfer enables higher metabolic activity and more than 40% volatile solids destruction in a retention time lesser than 4 d (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 involving the use of vertical shaft reactor is termed vertical shaft flotation. The thickened product dewaters to more than 30% solids with a relatively low polymer demand (approx 14 lb/t).

Advantages of the higher operating temperatures in the VSD-ATAD-Air and the VSD-ATAD-Oxygen systems include:

a. Reduced HRT (3–6 d) 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).
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 (<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 250–300° d is required to achieve reasonable volatile solids destruction (67,89).

8.2. Summary

There are seven primary variations of the aerobic digestion process:

a. Conventional aerobic digestion using air.
b. Conventional aerobic digestion using oxygen.
c. Auto thermal thermophilic aerobic digestion using air (ATAD-Air).
d. Auto thermal thermophilic aerobic digestion using oxygen (ATAD-Oxygen).
e. ATAD-Air process using vertical shaft reactor for aeration/digestion (VSD-ATAD-Air).
f. ATAD-Oxygen process using vertical shaft reactor for oxidation/digestion (VSD-ATAD-Oxygen).
g. 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 on 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:

a. Relatively simple operation with volatile solids reduction slightly less than anaerobic systems.
b. Low BOD concentrations in the effluent supernatant.
c. Production of an odorless, biologically stable end product.
d. Recovery of more of the sludge fertilizer value.

e. Reduced capital cost.

Reported disadvantages of all aerobic digestion processes include:

a. High power cost associated with aeration.
b. Higher polymer cost associated with product dewatering.
c. 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 2.5–10 ft in diameter, offer extremely high oxygen transfer efficiency, and extremely small foot print for construction. The HRT can be shortened to 3–6 d 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 since 1970 aerobic digestion has been used in northern cold climates. 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:

a. Solids content = 2191 lb/d.
b. Specific gravity = 1.05.
c. Solids concentration = 1%.
d. 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

1. First: calculate total quantity of raw sludge.

$$Q = \frac{W_s(100)}{(\text{specific gravity})(\%\ \text{solids})(8.34)} \tag{5}$$

where Q is the quantity of raw sludge production (gpd); W_s is the weight of solids in produced sludge (lb/d) = 2191 lb/d.

$$\begin{aligned} Q &= 2191\ (100)/(1.05)\ (1)\ (8.34) \\ &= 25{,}020\ (\text{gpd}). \end{aligned}$$

2. Second: select hydraulic detention time and calculate volume of digester.

$$V = (t)\ (Q) \tag{6}$$

where V is the volume of digester (gal); t is the hydraulic detention time = 15 d; Q is the quantity of raw sludge production = 25,020 gpd.

$$V = (15)\ (25{,}020)$$
$$= 375{,}300 \text{ gal.}$$

3. Third: check volatile solids loading.

$$L_{vs} = \frac{W_{vs}\ (7.48)}{V} < 0.1 - 0.2 \tag{7}$$

where L_{vs} is the volatile solids loading (lb/ft^3/d); W_{vs} is the weight of volatile solids production = 2191 (0.80) = 1753 lb/d; V is the volume of digester = 375,300 gal.

$$L_{vs} = 1753\ (7.48)/375{,}300$$
$$= 0.035 < 0.1.$$

4. Fourth: calculate solids retention time.
   a. Assume percent destruction of volatile solids = 50%.
   b. Calculate solids accumulation per day.

$$W_{ac} = W_s - W_s \text{ (volatile \%/100) (destroyed \%/100) (0.75)} \tag{8}$$

where W_{ac} is the solids accumulation (lb/d); W_s is the weight of solids in produced sludge = 2191 lb/d.

$$W_{ac} = 2191 - 2191\ (80/100)\ (50/100)\ 0.75$$
$$= 1534 \text{ lb/d.}$$

   c. Assume MISS in digester and calculate total digester capacity.

$$W_{dc} = (V)\ (\text{MLSS})\ (8.34)\ (10^{-6}) \tag{9}$$

where W_{dc} is the digester capacity (lb); MISS is the mixed liquor SS in digester = 12,000 mg/L; V is the volume of digester =375,300 gal.

$$W_{dc} = 375{,}300\ (12{,}000)\ (8.34)\ (10^{-6})$$
$$= 37{,}560 \text{ lb.}$$

   d. Calculate solids retention time.

$$SRT = \frac{W_{dc}}{W_{ac}} \tag{10}$$

where SRT is the solids retention time (d); W_{dc} is the digester capacity = 37,560 lb, W_{ac} is the solids accumulation 1534 lb/d.

$$\text{SRT} = 37{,}560/1534$$
$$= 24.5 \text{ d.}$$

5. Fifth: calculate sludge wasting schedule.
   Assume solids content in digested sludge is approx 2.5%.
   V_w is the total sludge in digester (lb) (100)/(specific gravity) (% solids) (8.34) (11)
   where V_w is the volume of sludge to be wasted each SRT (gal).

$$V_w = 37{,}560\ (100)/1.05\ (2.5\%)\ (8.34)$$
$$= 171{,}566 \text{ gal to be wasted every 24.5 d (about 7000 gal/d).}$$

6. Sixth: calculate oxygen requirement for bacterial growth.

Assume O_2 required per pound of volatile solids destroyed = 2 lb.

$$O_2 = (O_r)\ W_s\ (\text{volatile }\%/100)\ (\text{destroyed }\%/100) \tag{12}$$

where O_2 is the total oxygen required (lb/d); O_r is the oxygen required/lb of volatile solids destroyed = 2 lb; W_s is the weight of solids in produced sludge = 2191 lb/d.

$$\begin{aligned} O_2 &= 2\ (2191)\ (0.80)\ (0.50) \\ &= 1753 \text{ lb/d.} \end{aligned}$$

7. Seventh: calculate air requirement for mixing.
   a. Assume standard transfer efficiency = 8%.
   b. Assume constants α, β, and ρ.

$$\alpha = 0.9.$$
$$\beta = 0.9.$$
$$\rho = 1.$$

   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 is the operating transfer efficiency (%); ε_s is the standard transfer efficiency = 8%; α is the (K_{La} waste/K_{La} water) = 0.9; β is the (C_s waste/C_s water) = 0.9; ρ is the correction for altitude = 1; C_L is the minimum oxygen to be maintained in the digester = 2 mg/L, K_{La} is the oxygen transfer coefficient; T is the temperature = 25°C; $(C_s)_T$ is the oxygen saturation at the summer temperature = 8.2 mg/L.

$$\begin{aligned} \varepsilon_o &= 8\ [(8.2)\ (0.9)\ (1)-2]\ (0.9)\ (1.024)^{25-20}/9.20 \\ &= 4.7\% \end{aligned}$$

   e. Calculate air supply; check against a minimum of 20 cfm/1000 ft^3.

$$Q_{\text{air}} = O_2\ (7.48)\ (10^5)/(\varepsilon_o\%)\ (0.0176 \text{ lb } O_2\ /\text{ft}^3 \text{ air})\ 1440\ (\text{min/d})\ V \tag{14}$$

where Q_{air} is the air supply (cfm/1000 ft^3); O_2 is the total oxygen required = 1753 lb/d; ε_o is the operating transfer efficiency = 4.7%; V is the volume of the basin = 375,300 gal.

$$\begin{aligned} Q_{\text{air}} &= 1753\ (7.48)\ (10^5)/(4.7\%)\ (0.0176 \text{ lb } O_2\ /\text{ft}^3 \text{ air})\ 1440\ (\text{min/d})\ 375{,}300 \\ &= 29.3 \text{ cfm/1000 ft}^3 > 20. \end{aligned}$$

9.2. Example 2

A design engineer has determined that the following quantities of sludge will be produced at 0.5 MGD (22 L/s) contact stabilization plant:

a. Total daily solids generation = 1262 lb (572 kg).
b. Amount because of chemical sludge = 0.
c. Amount that will be volatile = 985 lb (447 kg).
d. Amount that will be nonvolatile = 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 more than 40% 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 8000 mg/L.
f. Expected thickened solids concentration for the stabilized sludge is 3% (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 6 offers a quick method for calculating the number of degree days required to achieve the 40% volatile solids reduction required. The result is 475° d. At a basin temperature of 50°F (10°C) then: 475° d/10° = 47.5 d.
Therefore, the volume of the aerobic digester must be adequate to provide 47.5 d sludge age to meet minimum volatile solids reduction during the winter. During the summer, the basin temperature will be 77°F (25°C): 25°C × 47.5 d sludge age = 1175° d. From Fig. 6, at 1175° d, there would be 49% volatile solids reduction.

b. Amount of volatile solids reduction.
For winter conditions, there would be a 40% volatile solids (VS) reduction. The actual pounds of solids reduced are:

$$985 \text{ lb VS/d} \times 0.4 = 394 \text{ lb VS reduced/d (179 kg/d).}$$

For summer conditions, there would be a 49% volatile solids reduction. The actual pounds of solids reduced are:

$$985 \text{ lb VS/d} \times 0.49 = 483 \text{ lb VS reduced/d (219 kg/d).}$$

c. Oxygen requirements.

Because nitrification is expected, provisions must be made to supply 2 lb of oxygen/ lb of volatile solids destroyed (2 kg O_2/kg volatile solids destroyed).
Winter conditions: 394 lb VS destroyed/d × 2 lbs O_2/lb VS destroyed = 788 lb O_2/d (358 kg/d).
Summer conditions: 483 lb VS destroyed × 2 lbs O_2/lb VS destroyed = 966 lb O_2/d (438 kg/d).
During summer conditions, a minimum of 1 mg/L oxygen residual must be provided.

d. Tank volume.
Sludge age in an aerobic digester can be defined as follows:
Sludge age is the 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 8000 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% of the thickened solids concentration, or 30,000 × 0.70 = 21,000 mg/L.
An average poundage of suspended solids in the supernatant W_{sss} can be approximated by the following equation:

$$W_{sss} = (\text{SS concentration in supernatant})\ (l\text{–}f)\ (8.34)\ (\text{influent flow})$$

where *f* is the fraction of influent flow into the aerobic digester that is retained; *1–f* is the fraction that leaves as supernatant.

The term *f* can be approximated by the following equation:

f is the (influent SS concentration/thickened SS concentration) (fraction of solids not destroyed).

For winter conditions, the fraction of solids not destroyed is: (1262 lb total solids–394 lb of solids reduced)/1262 lb total solids = 0.69.

Then, the term f for this example is:

$$(8000\ \text{mg/L}/30{,}000\ \text{mg/L}) \times 0.69 = 0.18$$

Therefore, 18% of the influent flow into the aerobic digester will be retained, and 82% will leave as supernatant. For a properly designed solids–liquid separator (less than 200 gpd/ft^2 = 8.16 m^3/d/m^2 overflow rate), the suspended solids concentration would be approx 300 mg/L. The influent flow can be found by dividing the influent solids load (1262 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 W_{ssw} from the aerobic digestion system can now be approximated from the following expression:

$$W_{ssw} = (\text{SS concentration in thickened sludge})\ (f)\ (8.34)\ (\text{influent flow}).$$

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 d minimum was required. From the values previously discussed:

47.5 d = (21,000 mg/L) (8.34) (*V* MG)/[300 mg/L (1–0.18) + (30,000) (0.18)] (8.34) (0.018915 MG)

Tank volume (*V*) = 0.233 MG (881 m^3).

Theoretical hydraulic detention time:

233,000 gal/18,915 gpd = 12.3 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% stand-by capacity as per state requirements).

e. Power requirement and tanks dimensions.

Select a depth of 12 ft (3.65 m). Because 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:

Area = V/depth.
Area = 233,000/7.48 (12).
= 2596 ft^2 (241 m^2).

The aerator manufacturer recommends that a minimum 10 hp unit (7.5 kW) will be required to mix the 12 ft (3.65 m) liquid depth. Each 10 hp unit (7.5 kW) could mix an area 40 × 40 ft^2 (12.1 × 12.1 m^2). After making some calculations, it is found that there is a need for two 10 hp (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. Figures 7 and 8 show 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 m^3/d/m^2). At an influent flow of 18,915 gal/d (71.5 m^3/d), the required surface area is:

Area = 18,915/(200) = 95 ft^2 (8.8 m^2).

Using a circular clarifier:

Diameter^2 = [95(4/3.14)] = 121.

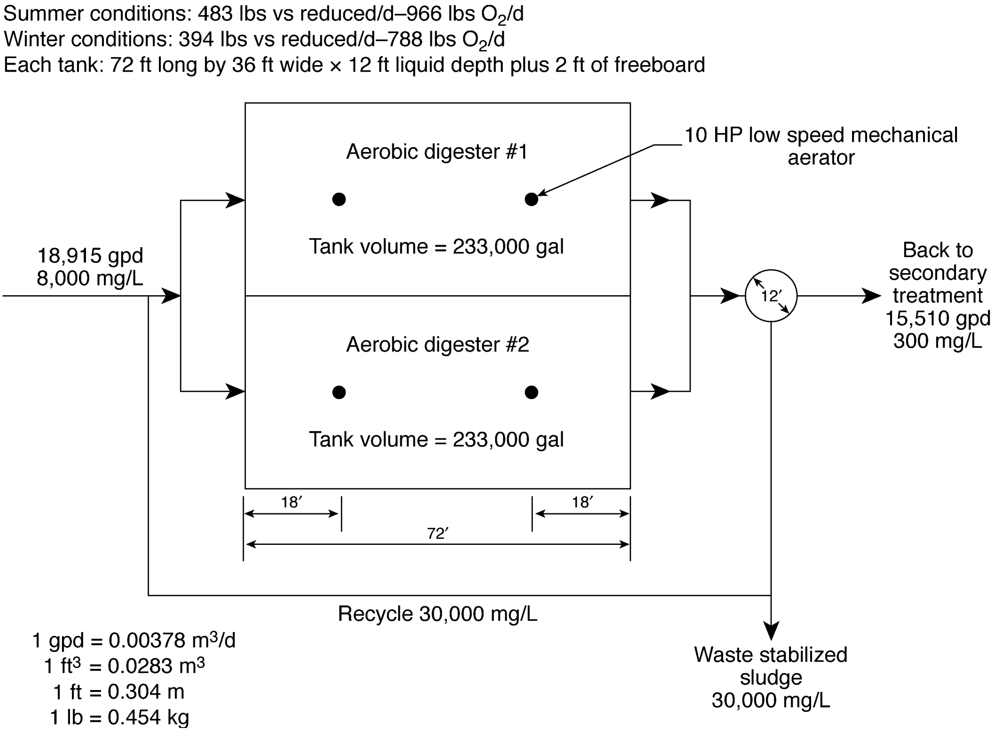

Fig. 7. Aerobic digestion design Example 2 (*Source:* US EPA) (9).

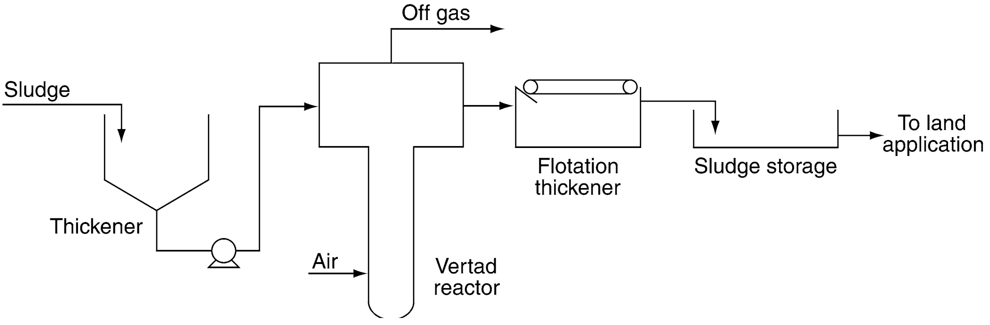

Fig. 8. Vertical shaft digestion (VSD) process system (VERTAD) (*Source:* NORAM Engineering and Constructors Ltd., Vancouver, Canada) (87,89).

Diameter = 11ft.

Select a 12-ft (3.7 m) diameter clarifier.

g. Supernatant flow.

It was previously calculated that 82% 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 gal/d (58.6 m^3/d), in addition to any precipitation.

NOMENCLATURE

α	(K_{La} waste/K_{La} water) = 0.9
β	(C_s waste/C_s water) = 0.9
C	Capital cost of process in 2006 USD
$C_{A\&E}$	Cost of administration and engineering in 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 (%)
ε_s	Standard transfer efficiency (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)
MISS	mixed liquor SS in digester (mg/L)
O_2	Total oxygen required (lb/d)
O_r	Oxygen required/lb of volatile solids destroyed = 2 lb
ρ	Correction for altitude = 1
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_{sss}	Weight of suspended solids in the supernatant (lb)
W_{ssw}	Weight of suspended solids wasted per day (lb)
W_{vs}	Weight of volatile solids production (lb/d)

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APPENDIX

United States Yearly Average Cost Index for Utilities[a]

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

[a] Extracted from US ACE (2000-Tables Revised 31 March 2003) *Civil Works Construction Cost Index System Manual,* no. 1110-2-1304, US Army Corps of Engineers, Washington, DC, PP 44 (PDF file is available on the Internet at http://www.nww.usace.army.mil/cost (72).

7
Lime Stabilization

Clint Williford, Wei-Yin Chen, Nazih K. Shammas, and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

Alkaline or lime stabilization is a very simple process. Its principal advantages over other stabilization processes include low cost and simplicity of operation. However, lime stabilization accomplished at a pH of 10.0–11.0, may allow odors to return during storage owing to pH decay. To eliminate this problem and reduce pathogen levels, addition of sufficient quantities of lime to raise and maintain the biosolids pH to 12.0 for 2 h is required. The lime-stabilized biosolids readily dewaters with mechanical equipment and is generally suitable for application on an agricultural land or disposal in a sanitary landfill.

No direct reduction of organic matter occurs in lime treatment. This has two important impacts. First, lime addition does not make biosolids chemically stable; if the pH drops less than 11.0, biological decomposition will resume, producing noxious odors. Second, the quantity of biosolids for disposal is not reduced, as it is a method of biological stabilization. On the contrary, the mass of dry biosolids is increased by the lime added and by the chemical precipitates that are derived from this addition. Thus, because of the increased volumes, the costs for transport and ultimate disposal is often more for lime-stabilized biosolids than for biosolids stabilized by other methods.

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Edited by: L. K. Wang et al. © The Humana Press Inc., Totowa, NJ

2. PROCESS DESCRIPTION

2.1. History

Lime has been traditionally used to reduce odor inconveniences such as from open pit privies and the graves of domestic animals. Lime has been used commonly in wastewater biosolids treatment to raise the pH in stressed anaerobic digesters and to condition biosolids prior to vacuum filtration. The original objective of lime conditioning was to improve biosolids dewatering, but it was observed that odors and pathogen levels were also reduced. In 1954, T. R. Komline filed a patent for a method of processing raw biosolids, in which heavy dosages of hydrated lime (6–12% of total dry solids) were added specifically to cancel or inhibit odors. However, only recently has lime addition been considered a major biosolids stabilization alternative.

Many studies describe the effectiveness of lime in reducing microbiological hazards in water and wastewater, but the bactericidal value of adding lime to biosolids has been noted only recently (1–4). A report of operations at the Allentown, Pennsylvania wastewater treatment plant states that lime conditioning of anaerobically digested biosolids to a pH of 10.2–11.0, and then vacuum filtering and storing the cake, destroyed all odors and pathogenic enteric bacteria (5). Kampelmacher and Jansen (6) reported similar experiences. Evans noted that lime addition to biosolids released ammonia and destroyed coliform bacteria, and that the biosolids cake was a good source of nitrogen and lime to the land (7).

Lime stabilization of raw biosolids has been conducted in the laboratories, as well as in full-scale plants. Farrell and others (8) reported that lime stabilization of primary biosolids reduced bacterial hazard to a negligible value, improved vacuum filter performance, and provided a satisfactory means of stabilizing biosolids prior to ultimate disposal. Paulsrud and Eikum (9) determined the lime dosage required to prevent odors from stored wastewater biosolids. Primary biological biosolids, septic tank biosolids, and different chemical biosolids were used in the study. An important finding was that lime dosages more than sufficient to raise the pH of the biosolids was required to prevent pH decay and the return of odor during storage. Laboratory and pilot scale work by Counts and Shuckrow (10) on lime stabilization, showed significant reduction in pathogen population and had a repelling odors when the biosolids pH was more than 12.0. Counts conducted growth studies on greenhouse and outdoor plots which indicated that the disposal of lime-stabilized domestic biosolids on cropland, would have no detrimental effect on plant growth and soil characteristics. Disposal of the lime-stabilized domestic biosolids at loading rates up to 100 t dry solids per hectare (224 T/ha) on green-house plots and 40 t dry solids/hectare (90 T/ha) on outdoor plots had no detrimental effect on plant growth and soil characteristics.

A full-scale lime stabilization facility was built as part of a 1 MGD (43.8 L/s) wastewater treatment plant in Lebanon, OH. The operation began in 1976. A case study of lime treatment and land application of biosolids from this plant is available along with a general economic comparison of lime stabilization with anaerobic digestion (11).

2.2. Current Status and Regulations

Biosolids are primarily organic materials produced during wastewater treatment, which may be put to beneficial use. Biosolids are used in home gardening, commercial agriculture, silviculture, greenways, recreational areas, and reclamation of drastically disturbed sites such as those subjected to surface mining. Biosolids are often rich in nutrients such as nitrogen and phosphorus, and contain valuable micro nutrients (12). The US Environmental Protection Agency's (US EPA) 40 CFR Part 503, *Standards for the Use and Disposal of Sewage Sludge* (the Part 503 Rule) requires that wastewater biosolids be processed (stabilized) before they can be beneficially used (13,14). Stabilization helps to minimize the potential for odor generation, destroys pathogens (disease causing organisms), and reduces the material's vector attraction potential. One method of stabilization is to add lime (an alkaline material), to raise the pH level to make conditions unfavorable for the growth of organisms (such as pathogens).

The Part 503 Rule (13) defines two types of biosolids with respect to pathogen reduction: class A (no detectable pathogens) and class B (a reduced level of pathogens). Both classes are safe, but additional requirements are necessary with class B materials. These requirements are detailed in the Part 503 rule and include such things as limiting public access to the site of application, limiting livestock grazing, and controlling crop harvesting schedules. Class A biosolids are not subject to these restrictions and can generally be used like any commercial fertilizer. Alkaline stabilization can achieve the minimum requirements for both classes A and B biosolids with respect to pathogens, depending on the amount of alkaline material added and other processes employed. Generally, lime stabilization meets the class B requirements when the pH of the mixture of wastewater solids and alkaline material is at 12, or after 2 h of contact.

Class A requirements can be achieved when the pH of the mixture is maintained at or more than 12 for at least 72 h, with a temperature of 52°C maintained for at least 12 h during this time. In one process, the mixture is air dried to over 50% solids after the 72 h period of elevated pH. Alternately, the process may be manipulated to maintain temperatures at or more than 70°C for 30 or more min, while maintaining the pH requirement of 12.0. This higher temperature can be achieved by an overdose of lime (i.e., adding more than is needed to reach a pH of 12.0), by using a supplemental heat source, or by using a combination of the both. Monitoring for fecal coliforms or *Salmonella* spp. is required prior to the release for use.

The Part 503 rule also allows for meeting class A pathogen reduction requirements through monitoring for pathogens before and after processing. For example, The third alternative for class A requires that the unprocessed wastewater biosolids be monitored for enteric viruses and helminth ova. The process is monitored for lime dosage and pH, and the final product must have no detectable levels of enteric viruses or helminth ova.

For more specific details on the requirements for achieving classes A or B, the reader is referred to the Part 503 rule (13,14). However, it must be mentioned here that the main requirements relate to the reduction of pathogens and vector attraction (15). Meeting class B requirements may be referred to as processes to significantly reduce pathogens (PSRP) lime stabilization; and meeting class A, as processes to further reduce pathogens (PFRP) lime pasteurization (13–17).

The product must meet requirements which are needed for secondary use. If the biosolids are to be disposed of at a permitted site, then class B requirements should be met; and for distribution to the public, class A requirements should also be met. Meeting the class B requirements requires equipment for biosolids feed and biosolids/lime

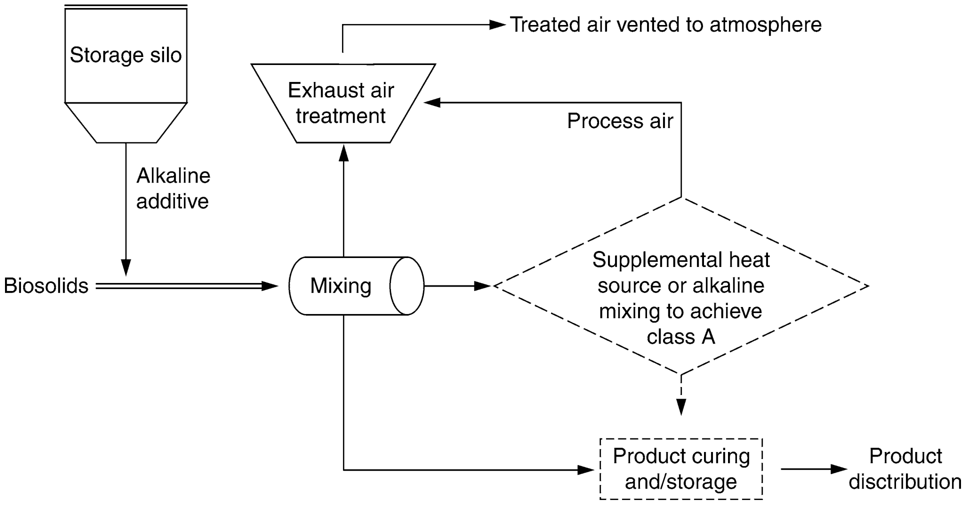

Fig. 1. Flow diagram of lime (alkaline) stabilization operation (*Source:* US EPA).

mixing (Fig. 1). The pH must be raised to 12.0 and held there for 2 h. Meeting class A requirements involves essentially the same feed and mixing equipment, with the additional thermal treatment and holding, or pasteurization vessel (Fig. 1). This will be described and further detailed with energy use and cost estimates in following sections.

Materials that may be used for alkaline stabilization include hydrated lime, quicklime (calcium oxide), fly ash, lime and cement kiln dust, and carbide lime. Quicklime is commonly used because it has high heat of hydrolysis (491 Btu) and can significantly enhance pathogen destruction. Fly ash, lime kiln dust, or cement kiln dust are often used for alkaline stabilization because of their availability and relatively low cost.

The alkaline-stabilized product is suitable for application in many situations, such as landscaping, agriculture, and mine reclamation. The product serves as a lime substitute, source of organic matter, and a speciality fertilizer. The addition of alkaline-stabilized biosolids results in more favorable conditions for vegetative growth by improving soil properties such as pH, texture, and water holding capacity. Appropriate applications depend on the needs of the soil and crops that will be grown and the pathogen classification. For example, a class B material would not be suitable for blending in a top soil mix intended for use in home landscaping but is suitable for agriculture, mine reclamation, and landfill cover, where the potential for contact with the public is lower and access can be restricted. Class A alkaline-stabilized biosolids are useful in agriculture and as a topsoil blend ingredient. Alkaline-stabilized biosolids provide pH adjustment, nutrients, and organic matter, reducing reliance on other fertilizers (12). Lime stabilized biosolids are also useful as daily landfill cover. They satisfy the federal requirement that landfills must be covered with soil or soil-like material at the end of each day (40 CFR 258). In most cases, lime stabilized biosolids are blended with other soil to achieve the proper consistency for daily cover.

As previously mentioned, lime stabilized biosolids are excellent for land reclamation in degraded areas, including acid mine spoils or mine tailings. Soil conditions at such sites are very unfavorable for vegetative growth often owing to acid content, lack of nutrients, elevated levels of heavy metals, and poor soil texture. Lime stabilized biosolids help to remedy these problems, making conditions more favorable for plant growth and reducing potential erosion. In addition, once a vegetative cover is established, the quality of mine drainage improves.

By the late 1970s, lime stabilization was being used to stabilize the biosolids from at least 27 municipal wastewater treatment plants in Connecticut. Average wastewater flows treated at these plants varied from 0.1 to 31 MGD (4.4–1358 L/s). Since then, the technology has been widely applied. In most of the plants, incinerators have been either wholly or partially abandoned. Although few chemical or bacterial data are available, qualitative observations indicate that treatment is satisfactory. Most of the communities have indicated that they will continue with lime stabilization (18).

Landfill burial may be used for disposal of lime-stabilized biosolids. However, lime-treated biosolids from eight plants in Connecticut are applied onto land. The biosolids are spread onto cornfields when application is compatible with crop cycles and weather conditions. The combination of lime, nutrients, and organic material is in demand, and has been used throughout the world to improve soil conditions and increase crop yields (12). Few inconveniences are caused by the practice. No odors eminated when piles have been opened for spreading of the biosolids. In Willimantic, Connecticut, lime-stabilized biosolids are mixed with leaves and grasses. After stockpiling, a portion of mixture is screened and distributed to local nurseries. The remainder is used as final cover for landfill.

2.3. *Applicability*

Where lime or another alkaline additive (for e.g., recycled kiln dust), is relatively inexpensive, alkaline stabilization is often the most cost-effective process for wastewater solids stabilization (12). This is particularly true where dependable markets for the alkaline product can be developed, such as in areas where alkaline materials are routinely applied to agricultural soils to maximize crop yields. Alkaline stabilization is practical at small wastewater treatment plants that store wastewater solids, to be transported later to larger facilities for further treatment. It is applicable for expansion of existing facilities or as a new facility to reduce odors and pathogens. The technology is especially useful at wastewater treatment facilities with flows that vary greatly because the process adjusts easily to changing flows. This adaptability also makes alkaline stabilization an appropriate choice as a secondary or backup stabilization method; because these facilities can be started and stopped relatively quickly and easily. Facilities can also be designed to handle either liquid or dewatered wastewater biosolids. It can be concluded that lime stabilization can be an effective alternative when there is a need to provide the following capabilities (18):

a. **Backup for existing stabilization facilities.** A lime stabilization system can be started (or stopped) quickly. Therefore, it can be used to supplement existing biosolids processing facilities when biosolids quantities exceed design levels, or to replace incineration during fuel shortages. Full biosolids flows can be lime-treated when existing facilities are out of service for cleaning or repair.

b. **Interim biosolids handling.** Lime stabilization systems have a comparatively low capital cost and, therefore, may be cost effective if there are plans to close the plant or process within a few years.
c. **Expansion of existing facilities or construction of new facilities to improve odor and pathogen control.** Lime stabilization is particularly applicable in small plants or when a plant is loaded seasonally.

In all cases, a suitable site for disposal or use of stabilized biosolids is required (19). In general, alkaline stabilization is not a proprietary process, which means no fee is required to be paid to a patent holder to use the process (20–22). However, several variations on the basic process are proprietary, such as:

a. BioFIX Process, marketed by Bio Gro Division of Wheelabrator Clean Water Systems, Inc., Millersville, MI (23).
b. RDP En-Vessel Pasteurization System, marketed by RDP Company (12).
c. N-Viro advanced alkaline stabilization with drying, marketed by N-Viro International Corporation, Toledo, OH (24).

2.4. *Theory of the Process*

Lime addition to biosolids reduces odor and pathogen levels by creating a high pH environment hostile to biological activity. Gases containing nitrogen and sulfur that are evolved during anaerobic decomposition of organic matter are the principal source of odor in biosolids (10,25). When lime is added, the microorganisms involved in this decomposition are strongly inhibited or destroyed in the highly alkaline environment. Similarly, pathogens are inactivated or destroyed by lime addition.

A high lime dosage in biosolids also affects the chemical and physical characteristics of biosolids (26,27). Although the complex chemical reactions between lime and biosolids are not well understood, it is likely that mild reactions, such as the splitting of complex molecules by hydrolysis, saponification, and acid neutralization, occur in the high pH environments created by lime stabilization (10). These reactions reduce the fertilizer value of the stabilized biosolids, improve its dewatering, and change the character of liquid sidestreams. The nature of these chemical changes is described in Section 4.4.

2.5. *Advantages and Disadvantages*

Alkaline stabilization offers several advantages, are as follows (12):

a. Consistency with the US EPA's national beneficial reuse policy results in a product suitable for a variety of uses and is easy to sell.
b. Simple technology which requires few special skills for reliable operation.
c. Easy to construct with commonly available parts.
d. Small area of land required.
e. Flexible operation which can be easily started and stopped.

Several possible disadvantages should be considered in evaluating this technology (12):

a. The resulting product is not suitable for use on all kinds of soils. For example, alkaline soils common in southwestern states will not benefit from the addition of a high pH material.
b. The volume of material to be managed and moved off-site is increased by about 15–50% in comparison with other stabilization techniques, such as digestion. This increased volume results in higher transportation cost when material is moved off-site.

c. There is potential for odor generation both at the processing and end use site.
d. There is a potential for dust production.
e. There is a potential for pathogen regrowth if the pH drops less than 9.5 while the material in store before use.
f. The nitrogen content in the final product is lower in comparison with several other biosolids products. During processing, nitrogen is converted to ammonia, which is lost to the atmosphere through volatilization. In addition, phosphorous available can be reduced through the formation of calcium phosphate.
g. There are fees associated with proprietary processes (class A stabilization).

2.6. Environmental Impacts

There are several potential environmental impacts associated with alkaline stabilization of wastewater solids (12). Odor problems may occur at the point of processing or use resulting from the release of ammonia and ammonia-related compounds and amines. These are generally considered as conveniences issues without long-term environmental impact. Handling of the material, such as loading, unloading, or spreading, all potentially cause release of ammonia and amines. The amount of ammonia released from the alkaline-stabilized product depends on the nitrogen content of the wastewater solids, and the pH and temperature achieved through the process. The extent of amine released will depend on the nature of the dewatering chemicals used.

In addition, small amounts of particulate matter may be emitted by the processing facility, but these are easily mitigated. Land application of any biosolid product can increase the concentrations of trace elements in the soil. Alkaline-stabilized biosolids help to create soil pH conditions in which metals are insoluble, minimizing plant uptake and movement of metals to groundwater. Soils which have a low pH will benefit greatly from the alkaline material and will be more fertile. Lime is usually low in metals and, when blended with wastewater solids, can improve the quality of the product with respect to metals (28).

3. DESIGN CRITERIA

Three fundamental design parameters must be considered in the design of a lime stabilization system: pH, contact time, and lime dosage. The primary objective of lime stabilization is to inhibit bacterial decomposition and inactivate pathogenic organisms. The extensive use of lime also provides an aesthetic benefit by masking noxious odors from decaying substances—facilitating its acceptance for biosolids treatment. Successful application requires information on conditions, necessary dose levels, and contact times for effective treatment.

The significant factor for effective lime treatment is the pH level, not just the dose of lime. As with most disinfection processes, the time of exposure (the extensive factor) is equally important as the pH (the intensive factor). Investigations, have established time, pH, and processing conditions as factors for producing satisfactory lime stabilization (8,10,11). Process performance is discussed in Section 4.

The design objective is to maintain pH more than 12.0 for about 2 h to ensure pathogen destruction, and to provide enough residual alkalinity so that the pH does not drop less than 11.0 for several days. This allows sufficient time for disposal or use without the possibility of renewed putrefaction. The recommended design criteria for accomplishing these objectives are (12,18):

Table 1
Lime Requirement to Attain pH 12.0 for 30 min

	Solids concentration (%)		Lime dosage (lb $Ca(OH)_2$/lb dry solids)		Average (pH)	
Sludge type	Range	Average	Range	Average	Initial	Final
Primary sludge[a]	3–6	4.3	0.06–0.17	0.12	6.7	12.7
Waste-activated sludge	1–1.5	1.3	0.21–0.43	0.30	7.1	12.6
Anaerobically digested mixed sludge	6–7	5.5	0.14–0.25	0.19	7.2	12.4
Septage	1–4.5	2.7	0.09–0.51	0.20	7.3	12.7

Source: US EPA.
[a] Includes some portion of waste-activated sludge.
1 lb/lb = 1 kg/kg.

a. Treat biosolids in the liquid state.
b. Bring the biosolids to pH 12.5 by lime addition and maintain pH more than 12.5 for 30 min (which keeps pH >12 for 2 h).
c. Common lime dosage required to achieve (b) is 0.25 lb lime/lb of wastewater solids at 20% solids.

The amount of lime required to stabilize biosolids is determined by the type of biosolids, its chemical composition, and the solids concentration. Table 1 summarizes the results of plant-scale tests at Lebanon, OH, and shows that lime additions ranging from 6 to 51% of the total dry solids in the biosolids were required to raise the pH to the levels indicated in the table (11). These lime dosages were sufficient to keep the biosolids pH more than 12.0 for 30 min. Primary biosolids required the lowest dosage, while the highest average dosages was required to raise the pH level of waste-activated sludge biosolids. Generally the results of studies conducted by Paulsrud and Eikum (9) agree with the Lebanon tests and are displayed in Table 2. Iron and alum biosolids required the highest dosage. Farrell and others (8) also found that alum additionally increased the lime requirement and suggested that part of the lime added to alum biosolids may be bound as a calcium–aluminum compound.

Figure 2 displays the general relationship between lime dosage and pH for a typical solids concentrations of municipal biosolids. Table 2 calculated from data on Fig. 2, shows that, to attain a particular pH, the required lime dose per unit mass of biosolids solids level is relatively constant. That is, lime requirements are more closely related to the total mass of biosolids solids, rather than the biosolids volume. Consequently, reduction in volume by thickening may have little or no effect on the amount of lime required, because the mass of biosolids solids is not changed.

Lime additions must be sufficient to ensure that the pH of biosolids does not drop below the desired level after prolonged storage. If insufficient lime is added, the pH will decay as the treated biosolids age (9–11). This phenomenon is displayed in Fig. 3. Notice that higher lime dosages not only raise the initial pH but, more importantly, prevent, or at least delay, the drop in pH levels. Consequently, in practice, lime doses must

Table 2
Lime Doses Required to Keep pH More Than 11.0 At least 14 d

Type of sludge	Lime dose (lb $Ca(OH)_2$/lb suspended solids)
Primary sludge	0.10–0.15
Activated sludge	0.30–0.50
Septage	0.10–0.30
Alum-sludge[a]	0.40–0.60
Alum-sludge[a] plus primary sludge[b]	0.25–0.40
Iron-sludge[a]	0.35–0.60

Source: US EPA.
[a] Precipitation of primary treated effluent.
[b] Equal proportions by weight of each type of sludge.
1 lb/lb = 1 kg/kg.

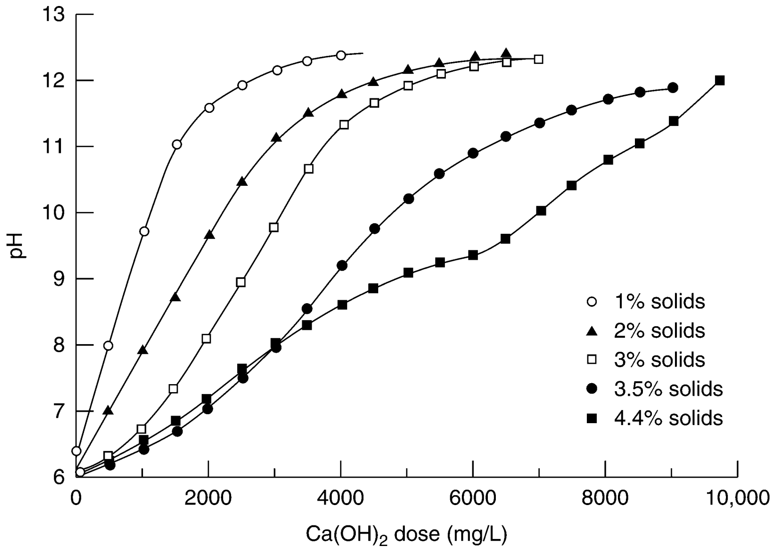

Fig. 2. Lime doses for raising the pH of primary/trickling filter biosolids mixture at different solids concentrations (*Source:* US EPA).

be greater than sufficient to raise the pH to the desired value. In most cases, significant pH decay will not occur if enough lime is added to raise the biosolids pH to 12.5 and maintain that value for at least 30 min (11).

Several mechanisms of pH decay have been proposed, and some have been documented (9,10). The initial pH decay results from the uptake of atmospheric CO_2, and slow reaction of hydroxyl ions with biosolids. The rate of pH decay is accelerated once the pH reaches a point at which bacterial action can resume production of organic acids through anaerobic microbial degradation.

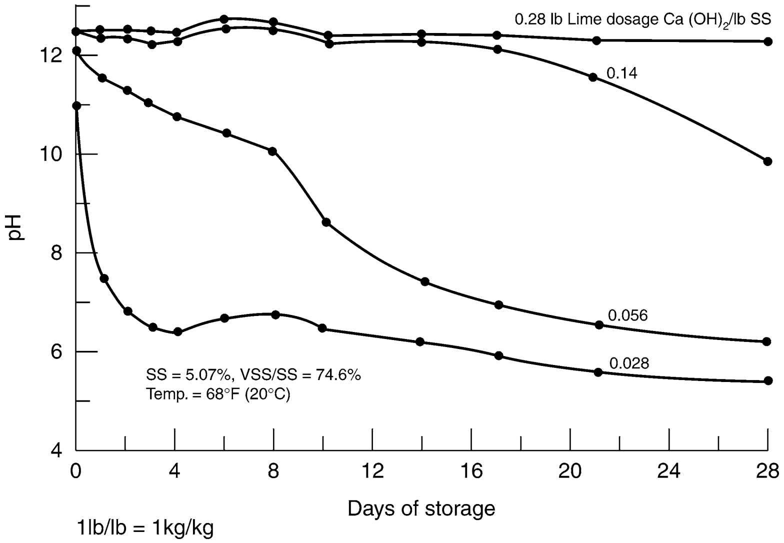

Fig. 3. Change in pH during storage of primary biosolids using different lime dosages (*Source:* US EPA).

There are many factors that must be considered in designing an alkaline stabilization facility for biosolids. The most critical are:

a. Percent solids of infeed wastewater solids.
b. Desired results (classes A vs B) which affect the amount of alkaline material and mixing time needed, which, in turn, impacts equipment size.
c. Source and volume of alkaline material to be used.
d. Odor control equipment at the processing facility.
e. Storage and curing areas.

The equipment necessary for alkaline stabilization is relatively easy to install and operate. Typical equipment includes the following (12):

a. Wastewater solids feed/conveyance mechanism.
b. Lime storage (silo, 1000 or 50 pound bags, and so on).
c. Lime transfer conveyor.
d. Mixer.
e. Air emission control equipment to minimize odor and dust.

Designing a facility to meet class A stabilization requirements may require additional lime storage to allow a larger lime dose, additional curing capacity, and/or the provision of supplemental heat. The end use of the material is another important factor when designing a biosolids management program, including alkaline stabilization. The resulting material is suitable for many uses, including agricultural application, mine reclamation and landfill cover. The amount of land that must be available differs with the use. Alkaline-stabilized biosolids are generally lower in nitrogen compared with other biosolids products because nitrogen is converted to ammonia during processing. The material contains approx 1–2% nitrogen, 1% phosphorus, negligible potassium, and nutrients of primary importance for vegetative growth. In agricultural application, alkaline-stabilized biosolids are often applied more for their pH adjusting characteristics at a typical application rate of 2–5 t/acre. Nutrients supplied to the crop are a secondary benefit. In reclamation applications, the material is often applied only once rather than annually or periodically as in agricultural applications. Therefore, the material is usually applied at a higher rate of between 60 and 100 dry t/acre. In landfill cover applications, the amount of material used is dictated more by regulatory requirements than nutrient content. Alkaline-stabilized biosolids can be used as daily cover, or to amend the final cover, in accordance with federal regulations which require material to enhance conditions for vegetative growth. For daily cover, a minimum of 6 in. of soil is required. Equivalent thickness of alkaline-stabilized material must be approved on a case by case basis. Typical application rates for incorporation into final cover material are similar to those at mine reclamation sites.

4. PROCESS PERFORMANCE

Alkaline stabilization is frequently selected by wastewater treatment facilities in regions where soils have a tendency to be acidic and where low cost alkaline materials are readily available. In areas where soils are acidic, the end product has greater value (12). Alkaline stabilization systems are generally quite reliable and flexible. The same equipment can often be used to produce either classes A or B biosolids with minor process modifications, such as a larger dose of alkaline material or the addition of supplemental heating (pasteurization). Lime stabilization reduces odor and odor production potential in biosolids, reduces pathogen levels, and alters dewatering, settling, and chemical characteristics of the biosolids. The nature and extent of the effects produced are described in the following paragraphs (18).

4.1. Deodorization

Lime treatment deodorizes biosolids by creating a high pH environment in the biosolids, thus eliminating or suppressing the growth of microorganisms that produce malodorous gases. In one laboratory study (10), the threshold odor number of raw mixed primary and trickling filter biosolids was 8000, while that of lime-treated biosolids ranged between 800 and 1300. The threshold odor number defines the greatest dilution of the sample with odor-free water to yield the least definite perceptible odor (29). Sufficient lime must be added to retard pH decay because odor generation will generally resume once the pH of the biosolids falls less than pH 11.0 (2,10). Hydrogen sulfide (H_2S), a malodorous gas present in dissolved form in biosolids, is a major cause of biosolids odors. Figure 4 shows that, as the pH of biosolids is raised, the fraction of total sulfide in the H_2S form decreases from about 50% at pH 7.0 to essentially zero at pH 9.0. Consequently, above this pH, there is no longer any H_2S odor (30).

In one example of odor reduction in a full-scale operation (11), odor was intense when septic raw biosolids were first pumped to the lime stabilization mixing tank. Odor intensity increased when diffused air was applied for mixing. When lime was added, the biosolids odor was masked by the odor of ammonia, which was stripped from the biosolids by the air which bubbled through the mixture. The ammonia odor was most intense with anaerobically digested biosolids and was strong enough to cause nasal irritation. As mixing proceeded, the treated biosolids acquired a musty, mucus-like odor.

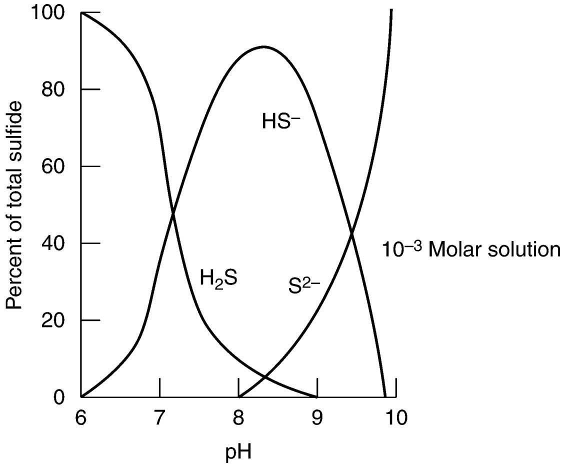

Fig. 4. Effect of pH on hydrogen sulfide/sulfide equilibrium (*Source:* US EPA).

In another work, it has been observed that two main compounds released during lime stabilization of biosolids are trimethylamine (TMA) and dimethyldisulfide. Factors affecting TMA and dimethyldisulfide production include the protein content of the feed, polymer usage, storage prior to liming, and enzyme-mediated transformations. TMA cannot apparently be eliminated under the anaerobic conditions normally present during solids processing. Management practices may reduce the TMA through limiting use of cationic polymers and limiting storage time. Future research will need to focus on reducing enzymatic conversion of proteins and altering precursors (31).

4.2. Pathogen Reduction

The most important aspect of lime stabilization concerns reduction of pathogens. As noted previously, there are two classes of treatment (A and B). The former involves heating to the specified temperatures (70°C for 30 min) to achieve pathogen reduction. The latter, class B, relies on elevation of pH, and is thus technically simpler, less rigorous and, thus more easily applied.

In the case of pathogen reduction by pH elevation, significant pathogen reductions can be achieved in biosolids that have been lime-treated to pH 12.0 (10,11). Table 3 lists bacteria levels measured during the full-scale study referred to earlier and shows that lime stabilization of raw biosolids reduced total coliform, fecal coliform, and fecal streptococci concentrations by more than 99.9%. The numbers of *Salmonella* and *Pseudomonas aeruginosa* were reduced lower than the level of detection. Table 3 also shows that pathogen concentrations in lime-stabilized biosolids ranged from 10 to 1000 times less than those in anaerobically digested biosolids from the same plant (12).

Information on virus destruction in biosolids by lime stabilization is scant. There are numerous investigations on removal of viruses from wastewater by lime flocculation but little on destruction of viruses by elevated pH. A study by Berg (32), measured the structure of a polio virus in water by pH adjustment alone, and indicated very rapid destruction at more than pH 11.0. Similar effects would be expected for other animal

Table 3
Bacteria in Different Types of Biosolids

	Bacterial density (number/l00 mL)				
Sludge type	Total coliform[a]	Fecal coliform[a]	Fecal *streptococci*	*Salmonella*	*Pseudomonas aeruginosa*
Raw					
Primary	2.9×10^9	8.3×10^8	3.9×10^7	62	195
Waste activated	8.3×10^8	2.7×10^7	1×10^7	6	5.5×10^3
Septage	2.9×10^8	1.5×10^7	6.7×10^5	6	754
Anaerobically digested					
Mixed primary and waste activated	2.8×10^7	1.5×10^6	2.7×10^5	6	42
Lime stabilized[b]					
Primary	1.2×10^5	5.9×10^3	1.6×10^4	<3	<3
Waste activated	2.2×10^5	1.6×10^4	6.8×10^3	<3	13
Septage	2.1×10^3	265	665	<3	<3
Anaerobically digested	18	18	8.6×10^3	<3	<3

Source: US EPA.

[a] Millipore filter technique used for waste-activated sludge and septage. MPN technique used for other sludges.

[b] To pH ≥ 12.0.

viruses. Qualitative observation under a microscope has shown substantial survival of higher organisms, such as hookworms, amoebic cysts, and *Ascaris* ova, after contact times of 24 h at high pH (8). It is not known whether long-term contact would eventually destroy these organisms. A detailed discussion of biosolids disinfection is contained in other chapters of this book (33).

Alkaline stabilization can result in an exceptional quality product when it meets class A pathogen reduction, vector attraction reduction and, the highest standards for metal concentrations (12). This higher level of processing can result in a more valuable product because there are no restrictions on end use. The high level of disinfection achieved in class A products makes them easier to handle and apply. For example, if a farmer purchases an exceptional quality product, he will not have to restrict access or limit grazing and harvesting times.

4.3. Dewatering and Settling Characteristics

Lime has been used extensively as a conditioning agent to improve the dewatering of biosolids. Trubnick and Mueller (34), presented detailed procedures to be followed in conditioning biosolids for filtration, using lime with and without ferric chloride. Sontheimer (35), described the improvements in biosolids filterability produced by lime addition.

The addition of lime has been shown to improve the filterability of alum and iron primary biosolids (8). Specific resistance was reduced by a factor of approx 4, and filter yield was increased by a factor of two when lime conditioning was used. According to the US EPA (10), the effect of lime treatment on the filterability of primary biosolids and trickling filter biosolids but could not detect any consistent trend.

Several investigators have examined the impact of lime stabilization on sand bed drying of biosolids (8,10,11). Lime additions to raw biosolids increased the rate of drying at least initially and, in one study, produced a drier final cake. However, lime-treated primary biosolids did not dry as fast as either lime-treated or untreated anaerobically digested biosolids (11). The settling of lime-stabilized primary and mixed biosolids was enhanced in one study (10), indicating that gravity thickening after lime treatment may be used to reduce the volume of biosolids to be dewatered. Regarding physical characteristics of the stabilized biosolids, the mixer should break the biosolids up and mix with lime to produce a granular consistency with 95% less than 0.5 in. Excessive energy input should be avoided to prevent formation of a thixotropic paste. The granular material is more easily spread by conventional farm equipment.

4.4. Chemical Characteristics

Lime stabilization causes chemical changes in the biosolids. The nature of these changes is illustrated in Tables 4 and 5, which compile chemical data from two studies (10,11). The general effect of lime addition is a reduction in component concentration. This is caused by both dilution with lime slurry and loss of some volatile biosolids components to the atmosphere.

Lime-stabilized biosolids have lower concentrations of soluble phosphate, ammonia nitrogen, and total Kjeldahl nitrogen compared with anaerobically digested biosolids from the same plant, as shown in Table 4. These lower nutrient levels reduce the agricultural value of the biosolids but, assuming nitrogen level are at the rate applicable to biosolids, it would allow more biosolids to be applied per hectare of land. A reduction in the soluble (filterable) phosphate concentration is caused by the reaction between lime and dissolved orthophosphate to form calcium-phosphate precipitate. For this reason, residual phosphate in the supernatant/filtrate after lime treatment is believed to be largely organic in nature (10). Nitrogen levels can be reduced during lime stabilization if gaseous ammonia is stripped during air mixing of the treated biosolids. As the pH of the biosolids increases from near neutral to 12.0, the predominant form of ammonia shifts from the ammonium ion (NH_4^+) to dissolved ammonia gas (NH_3). Some of this gas is carried off by the air bubbled through the biosolids for mixing.

A direct result of adding lime to biosolids is that the total alkalinity will rise to a high value. This can affect the suitability of the treated biosolids for land application. The input can be positive or negative, depending on soil conditions at the application site. Data in Table 4 indicates the magnitude of change in alkalinity.

Biochemical oxygen demand, chemical oxygen demand, and total organic carbon concentration increase in the liquid fraction of wastewater biosolids when lime is added (10,11). Organic matter is dissolved in the high pH environment. Possible reactions include saponification of fats and oils, hydrolysis and dissolution of proteins, and decomposition of proteins to form methanol (10).

Lime stabilization usually does not produce the substantial reductions in volatile matter associated with anaerobic and aerobic biosolids digestion. However, volatile solids concentration decreased by 10–35% after lime additions in the plant-scale studies at Lebanon (11), as shown in Table 4. The reductions in total solids concentration after lime stabilization, as displayed in Table 5, are greater than can be accounted for simply

Table 4
Chemical Composition of Biosolids Before and After Lime Stabilization

	Average concentration (mg/L)								
Sludge type	Alkalinity	Total COD	Soluble COD	Total phosphate	Soluble phosphate	Total Kjeldahl nitrogen	Ammonia nitrogen	Total suspended solids	Volatile suspended solids
Primary									
Before lime addition	1885	54,146	3046	350	69	1656	223	48,700	36,100
After lime addition	4313	41,180	3556	283	36	1374	145	38,370	23,480
Waste activated									
Before lime addition	1265	12,810	1043	218	85	711	51	12,350	10,000
After lime addition	5000	14,700	1618	263	25	1034	64	10,700	7136
Anaerobically disgested mixed sludge									
Before lime addition	3593	66,372	1011	580	15	2731	709	61,140	33,316
After lime addition	8467	58,670	1809	381	2.9	1780	494	66,350	26,375
Septage									
Before lime addition	1897	24,940	1223	172	25	820	92	21,120	12,600
After lime addition	3475	17,520	1537	134	2.4	597	110	23,190	11,390

Source: US EPA.

Table 5
Chemical Composition of Biosolids and Supernatant Before and After Lime Stabilization

Parameter	Primary sludge	Trickling filter humus	Mixed sludge
Whole sludge			
pH			
Before lime addition	6.0	6.3	6.1
After lime addition	12.1	12.3	12.0
Total solids (wt%)			
Before lime treatment	3.6	3.0	3.6
After lime treatment	3.2	2.7	3.3
Total alkalinity (mg/L as $CaCO_3$)			
Before lime addition	1141	1151	1213
After lime addition	6920	6240	5760
Ammonia nitrogen (mg N/L)			
Before lime addition	211	274	192
After lime addition	91	148	87
Organic nitrogen (mg N/L)			
Before lime addition	1066	1179	1231
After lime addition	1146	995	1099
Nitrate nitrogen (mg N/L)			
Before lime addition	3	7	16
After lime addition	25	22	31
Total phosphate (mg P/L)			
Before lime addition	342	305	468
After lime addition	302	235	337
Filterable phosphate (mg P/L)			
Before lime addition	92	96	80
After lime addition	32	17	31
Supernatant			
TOC (mg/L)			
Before lime addition	1000	917	1175
After lime addition	2083	1883	2250
BOD (mg/L)			
Before lime addition	1120	964	1137
After lime addition	1875	1981	2102
Threshold odor number[b]			
Before lime addition	4889	5333	933
After lime addition	467	333	67
Total solids (wt%)			
Before lime addition	0.1	0.1	0.2
After lime addition	0.6	0.5	0.7

Source: US EPA.

[a] Values in this table are averages of three tests for each sludge type.

[b] The greatest dilution with odor-free water to yield the least perceptible odor.

by dilution with lime slurry (10). It may be assumed that the lime interfered with the volatile solids analysis. However, reactions between lime and nitrogenous organic matter may cause a loss of sludges biosolids. Hydrolysis of proteins and destruction of amino acids are known to occur by reaction with strong bases. Volatile substances such as ammonia, water, and low molecular weight amines or other volatile organics may possibly be formed and lost to the atmosphere.

5. PROCESS DESIGN

A lime stabilization operation is divided into two operations: lime handling and biosolids mixing. Lime handling comprises facilities for receiving, storing, transporting, feeding, and "slurrying" of the lime. The biosolids mixing operation consists of a holding tank provided with mixing. A discussion of design considerations for these two operations is as follows.

5.1. Design of Lime Handling Facilities

Lime, in its various forms, is the principal and the lowest cost alkali used in industry and wastewater treatment. As a result, a substantial body of knowledge has evolved concerning the most efficient handling of lime. Only the basic elements of lime system design are described in this chapter. Detailed information is contained in several references that focus on the selection, handling, and use of lime (19,36–41).

5.1.1. Lime Characteristics

Lime is a general term applied to several chemical compounds that share the common characteristic of being highly alkaline. The two forms commercially available are quicklime (CaO) and hydrated lime [$Ca(OH)_2$]. The characteristics of these two chemicals are summarized in Table 6. Lime is a caustic material and can cause severe injury to tissue, particularly to eyes. Equipment must be designed with safe handling in mind; eyewash fountains and safety showers should be provided, and operating procedures should mandate use of proper handling procedures and protective clothing.

Quicklime is derived from limestone by a high temperature calcination process. It consists primarily of the oxides of calcium and magnesium. The grade of quicklime most commonly used in wastewater treatment contains 85–90% CaO (18). Quicklime is rarely applied directly (that is, in a dry condition) to the biosolids. First it is converted to hydrated lime by reaction with water in an exothermic reaction called slaking.

$$CaO + H_2O \rightarrow Ca(OH)_2 + \text{Heat} \tag{1}$$

During slaking, the generally coarse CaO particles are ruptured, splitting into microparticles of $Ca(OH)_2$. These smaller particles have a large total surface area and are highly reactive. The slaking reaction is carried out under closely controlled conditions to promote maximum lime reactivity.

If the lime manufacturer does the slaking, hydrated lime is delivered to the wastewater treatment plant. The manufacturer adds only enough water for hydration, producing a dry $Ca(OH)_2$ powder. At the waste treatment plant, the powder is then slurried with more water before mixing with biosolids. Alternatively, slaking may be carried out at the wastewater treatment plant; the delivered product is, therefore, quicklime. In this case, the lime is slaked, then diluted (if necessary) prior to process application.

Direct addition of dry quicklime to biosolids without the use of a separate slaker, is practiced in Denmark and in at least ten Swedish treatment plants (18). Potential advantages are the elimination of slaking equipment and the generation of heat, which can improve pathogen reduction and speed dewatering through evaporation. In one case,

Table 6
Characteristics of Quick Lime and Hydrated Lime

Common name/ formula	Available forms	Containers and requirements	Appearances and properties	Weight (1b/ft^3[bulk density])	Commercial strength	Solubility in water
Quick lime/CaO	Pebble Crushed Lump Ground Pulverized	80–100 lb moisture-proof bag, wooden barrels and car-loads. Store dry: maximum 60 d in tight container 3 mo in moisture-proof bag	White (light gray, tan) lumps to powder. Unstable caustic irritant slakes to hydroxide slurry evolving heat (490 btu/lb). Air slakes to $CaCO_3$. Saturated volume approx pH 12.5	55–75; to calculate hopper capacity– use 55; specify gravity 3.2–3.4	70–96% CaO (less than 88%) can be poor quality	Reacts to form $Ca(OH)_2$ each lb of quick lime will form 1.16 to 1.32 lb of $Ca(OH)_2$ with 2–12% grit, depending on purity
Hydrated lime/ $Ca(OH)_2$	Powder (passes 200 mesh)	50 lb bags, 100 lb barrels and car-loads. Store dry; maximum 1 yr	White, 200–400 mesh powder free of lumps caustic dusty irritant absorbs H_2O and CO_2 from air to form $Ca(HCO_3)_2$. Saturated solution approx pH 12.4	25–40, to calculate hopper capacity use 30, specify gravity 2.3–2.4	$Ca(OH)_2$ 82–98%; CaO 62–74% (Standard 70%)	10 lb/1000 gal at 70°F 5.6 lb/1000 gal at 175°F

Source: US EPA.

direct additions of dry quicklime were made to raise biosolids pH more than 13.0 and bring the temperature to 176°F (80°C). *Salmonella* and intestinal parasites were killed within 2 h. Heat generated by slaking of quicklime does not raise temperature significantly unless the biosolids are dewatered and the lime dose is high on the order of 400–800 lb/t dry solids (200–400 kg/T). Here t = 2000 lb, and T = 1000 kg.

The decision whether to purchase quicklime or hydrated lime in a particular situation is influenced by a number of factors such as the size of treatment facility, material cost, and storage requirements. The cost of hydrated lime is about 30% more than the cost of quicklime with equivalent calcium oxide content. The difference is owing to the higher production and transportation costs for hydrated lime. Nevertheless, hydrated lime is preferred for small-scale operations mainly because its use eliminates the labor and equipment required for slaking. Hydrated lime is also more stable and therefore is easier to handle and to store. When lime use exceeds 3–4 T/d (3000–4000 kg/d), quicklime should be considered because it is economical (36). Selection of the type of lime to be used should be based on a detailed economic analysis, taking into account all the unique factors of the particular application.

Both quicklime and hydrated lime react spontaneously with atmospheric CO_2.

$$CaO + CO_2 \rightarrow CaCO_3 \tag{2}$$

$$Ca(OH)_2 + CO_2 \rightarrow CaCO_3 + H_2O \tag{3}$$

In addition, quicklime can be slaked by the water vapor in the air.

These reactions cause two problems:

a. Lime quality is degraded because the reaction product, $CaCO_3$, is ineffective in raising pH.
b. The partial reaction with CO_2, and in the case of CaO, with water vapor, causes caking. This interferes with lime slaking and feeding.

Thus, lime storage, slaking, and feeding equipment should be sealed to as great a degree as possible to prevent contact of lime with atmospheric CO_2 and water vapor.

5.1.2. Delivery and Storage of Lime

Lime can be delivered either in bags or in bulk. The choice depends mostly on the rate of chemical use at the treatment plant. Bagged lime costs about 20% more than bulk lime, hut it is generally preferred where daily requirements are less than 1000–1500 pounds of lime/d (36). At this small scale, handling and storage of bagged lime is relatively simple, involving manual labor or simple mechanical aids. As the scale of operation increases, it becomes more efficient and economical to use bulk lime, which can be delivered in large quantities, transported in mechanical or pneumatic conveying systems, and stored in weather-tight bins or silos.

Bagged lime must be stored under cover to prevent rain from wetting the bags. Proper handling is especially important when quicklime is used, because it is highly reactive with water, producing heat and can bloat to cause the bags to burst. Because heat can be generated during accidental slaking of quicklime, bags should never be stored close to combustible materials.

Hydrated lime may be stored under dry conditions for periods up to a year without serious deterioration by reaction with atmospheric CO_2 (recarbonation). Quicklime deteriorates more rapidly. Under good storage conditions, with multiwall moisture-proofed

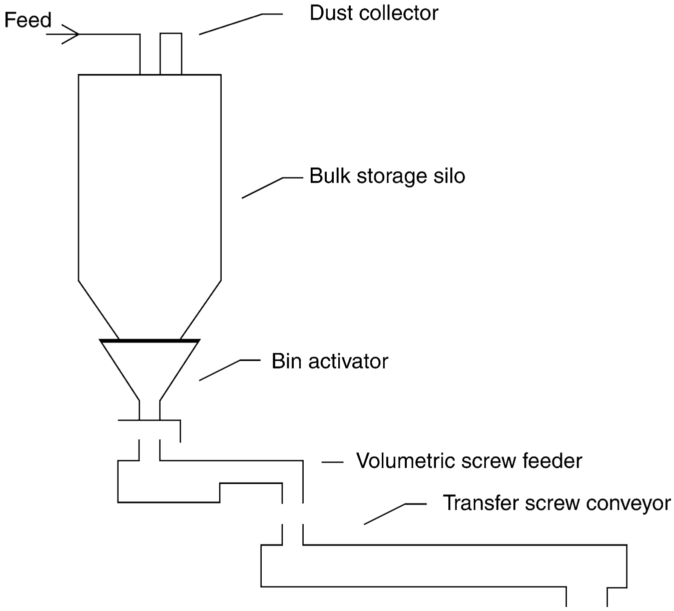

Fig. 5. Lime storage and feed equipment.

bags, quicklime may be held as long as 6 mo, but in general should not be stored for more than 3 mo (36,40,41).

Bulk quicklime and hydrated lime are delivered by trucks. These are equipped with blowers and 4 in. quick connect hoses to feed the lime to the silo. The silo should be equipped with an exhaust fan (at least 1000 cfm capacity) and a dust collector (preferably a mechanical shaker type) with approx 250 ft^2 of cloth area. The silos may be of conventional steel or concrete silos or bins. The storage facilities must be airtight to prevent slaking and recarbonation. Pebble quicklime is free-flowing and will discharge readily from storage bins if the hopper bottoms have a minimum slope of 60° from the horizontal. Pulverized quicklime and especially hydrated lime have a tendency to arch and therefore require some type of mechanical or aeration intrusion to ensure uniform discharge from storage bins. Detailed descriptions of the various types of flow-aiding devices can be found elsewhere (36,42).

The size of storage facilities should be determined taking into consideration daily lime demand, type and reliability of delivery, future chemical requirements, and an allowance for flexibility and expansion. At the least storage should be provided to supply a 7 d lime demand; however, sufficient storage to supply lime for 2–3 wk is desirable. In any case, the total storage volume should be at least 50% greater than the capacity of the delivery railcar or truck to ensure adequate lime supply between shipments (36). A recommended storage capacity based on truck delivery calls for a minimum of 1–0.5 truckloads to a maximum of 1 mo capacity. Figure 5 shows a recommended storage silo and feed system (16). Finally, the lime delivered into the silo will be aerated and unsettled, having a bulk density less than published values. Table 7 provides bulk densities for the main forms of lime, in comparison to the published values (16). Generally,

Table 7
Bulk Density for Lime–Published vs Delivered Values

Type	Published density (lb/ft^3)	Density for estimating silo size (lb/ft^3)
Pebble lime	55–60	45
Pulverized lime	70–75	60
Hydrated lime	35	25

the most economical silos have a diameter of 12 ft. A recommended height would be 16 ft, with a 60° conical bottom.

5.1.3. Lime Feeding

Lime is nearly always delivered to the biosolids mixing vessel as a $Ca(OH)_2$ slurry (milk-of-lime). This facilitates transport to the point of application and improves lime dispersion and reaction efficiency. The exact series of steps through which dry lime is wetted and introduced to the biosolids varies according to factors such as the scale of the operation, the type of lime purchased, and the method of storage. The following paragraphs outline the basic lime-feeding schemes. The discussion is largely derived from a bulletin published by the National Lime Association (36), which should be referred to for more detail.

In small treatment plants where bagged hydrated lime is purchased, the dry chemical is simply mixed with water in a batch tank and metered to the biosolids mixing tank as required. Solutions of lime are not corrosive, so that an unlined steel tank is sufficient for mixing and storage of the slurry. Hydrated lime is fed as a 6–18% $Ca(OH)_2$ slurry by weight, the percent depending on the application and on operator preference. The milk-of-lime can be discharged to the biosolids in one batch or metered continuously to the basin through a solution feeder.

In larger operations, where hydrated lime is stored in bulk, more automated mixing and feeding scheme is appropriate. These systems begin with a vibrating bin activator bolted to the bottom of the silo. These should have a maintenance gate to block off the lime during maintenance of downstream equipment. The lime then flows to a dry chemical feeder for continuous delivery of a measured amount of dry lime to a dilution tank. The feeder is often positioned directly at the base of the bulk storage bin to minimize dry lime transport distance.

Two general types of automated dry feeders are available:

a. Volumetric feeders, which deliver a constant, preset volume of chemical in a unit of time, regardless of changes in material density. Two main types are available—rotary valve and volumetric screw feeder. The latter experiences less abrasion, require less maintenance, and is more reliable than the rotary valve feeders (16).
b. Gravimetric feeders, which discharge a constant weight of chemical in a unit of time.

Gravimetric feeders cost roughly twice as much as volumetric feeders with an equivalent capacity and require more maintenance, but they are more accurate. Most manufacturers of gravimetric feeders will guarantee a minimum in accuracy of 1% percent, by weight, of the set rate. Volumetric feeders, on the other hand, may have an error of

30% by weight, owing to the varying bulk density of hydrated lime. Gravimetric feeders are preferred because of their greater accuracy and dependability, but the less expensive volumetric type may be sufficient when limited funds are available, when greater chemical feeding accuracy is not required, or when a reduced degree of maintenance is desirable. A volumetric screw feeder can convey the lime up to 2–3 ft. For greater distances, a transfer screw conveyor should be used (16).

Dry hydrated lime is delivered to a dilution tank that is often fitted directly onto the feeder. Compressed air, water jets, or impeller type mixers agitate the tank. The lime slurry is then transferred to the biosolids mixing basins. This transfer operation is the most troublesome operation in the lime handling process. The milk-of-lime reacts with atmospheric CO_2 or carbonates in the dilution water to form hard, tenacious $CaCO_3$ scales, which, with time, can plug the transfer line. Because the magnitude of this problem is in direct proportion to the distance over which the slurry must be transferred, lime feeder facilities should be located as close as possible to the lime/biosolids mixing tanks (18). Pumping of the lime slurry should be avoided (if possible, gravity transfer should be used), and all apparatus should be accessible for cleaning. Scaling in lime slurry systems has been prevented through the use of a chemical additive that interferes with crystal formation.

Feeding of quicklime is similar to for hydrated lime, except that there is an additional step, slaking, in which the quicklime reacts spontaneously with water to form hydrated lime. Bagged quicklime can be slaked in batches by simply mixing one part quicklime with two to three parts water in a steel trough while blending with a hoe. Proportions should be adjusted so that the heat of the reaction maintains the temperature of the reacting mass near 200°F (93°C). The resulting thin paste should be held for 30 min after mixing to complete hydration. Manually operated batch slaking is a potentially hazardous operation and should be avoided if possible. Uneven distribution of water can produce explosive boiling and splattering of lime slurry. Use of protective equipment should be mandatory. For small plants, the advantage of a using the lower-priced quicklime is smaller, because lime consumption is smaller. Use of slaked lime is safer, simpler, and requires less labor.

Continuous slaking is accomplished in automated machines that also dilute and degrit the lime slurry. Several types of continuous shakers are available. They vary mainly in the proportion of lime to water mixed initially. A volumetric or gravimetric dry chemical feeder is used to measure quicklime as it is moved from bulk storage to the shaker. Since quicklime is available in a wide range of particle sizes, it is important to match the dry feeder with the type of quicklime to be used in the particular application.

5.2. Biosolids-Lime Mixing Tank Design

A tank must be provided for mixing raw biosolids with lime slurry and then holding the mixture for a minimum contact time. It is recommended that all lime stabilization facilities should include a tank large enough to hold the lime biosolids mixture for 30 min. The pH of the reacted mixture should exceed 12.5 during this period. The following paragraphs discuss two aspects of mixing tank design—tank sizing and tank mixing.

5.2.1. Tank Sizing Considerations

Mixing tanks can be operated in either batch or continuous flow process modes. In the batch mode, the tank is filled with biosolids, and then sufficient lime is added to maintain the pH of the biosolids–lime mixture at more than 12.5 for the next 30 min. After this minimum contact time, the stabilized biosolids can be transferred to dewatering facilities or to either tank trucks or a pipeline for land application. Once the holding tank is emptied, the cycle begins again.

In the continuous flow mode, the pH and volume of biosolids in the holding tank are held constant. Entering raw biosolids displace an equal volume of treated biosolids. Lime is added continuously, in proportion to the flow of incoming raw biosolids, and thus, the holding time will vary. The lime dose must be sufficient to keep the contents of the tank at a pH of 12.5. Often the daily cycle of biosolids production does not match the pattern of biosolids disposal. In this case, a system could be operated on a semicontinuous basis, where the quantity of biosolids in the tank fluctuates through the day. Here the treatment tank would be used as a buffer between biosolids production and disposal.

It is most common to operate lime stabilization systems in the batch flow mode. Batch operations are very simple and are well suited for small-scale, manually operated systems. When adequate capacity is provided, the mixing tanks can also be used to gravity thicken the lime-treated biosolids before disposal. In very small treatment plants, tank capacity should be adequate to treat the maximum day biosolids production in one batch. This is because small plants are generally operated only during the day, and it is usually desirable to stabilize the entire day’s biosolids in one batch. Larger plants are more likely to be manned round-the-clock. Because biosolids can be processed over the whole day, stabilization tanks can be relatively smaller.

Continuous-flow stabilization systems require automated control of lime feeding and therefore are usually not cost-effective for small-scale operations. The primary advantage of continuous-flow systems over batch systems is that a smaller tank size may be possible. Capacity does not have to be provided for storage of biosolids between batches. Instead, the mixing tank must only be large enough to ensure that all biosolids particles are held at high pH for a contact time sufficient to destroy odor and disease-producing organisms.

The most important design parameter for a continuous flow, well-mixed reactor is the nominal detention time (defined as tank volume divided by volumetric input flow rate). Unlike a batch tank, where contact time of all particles is the same, some particles in a well mixed, continuously fed tank escape after relatively short contact. Thus, 30 min of pH at 12.5 in a batch mixer might not be the same as 30 min residence time in a well-mixed, continuously fed reactor.

In making a recommendation for detention time, the nature of the treatment that occurs must be considered. Unlike some treatments, such as irradiation, the treatment does not stop after the treated biosolids leaves the vessel. If pH is 12.5 as the biosolids leave the mixing tank, it remains at this pH after leaving. Consequently, a 30 min detention time in a continuously fed, well-mixed reactor is adequate, provided the pH is measured in an exit line. If pH of the limed biosolids appears to fall too rapidly upon standing, it is a simple matter to move the pH sensor and to control lime feed rate to a position further downstream.

Thickening of raw biosolids before lime addition will reduce the mixing tank capacity requirement in direct proportion to the reduction in biosolids volume. However, the lime requirement will be reduced only slightly by prethickening, because most of the lime demand is associated with the solids (9), and total solids mass is not changed by thickening.

5.2.2. Tank Mixing

Lime/biosolids mixtures can be mixed with either diffused air or mechanical mixers. The agitation should be great enough to keep solids suspended and to distribute the lime slurry evenly and rapidly. Both diffused air and mechanical systems can provide adequate mixing, although the former has been more commonly used in pilot studies and full-scale operations. In addition to their mixing function, sparger air systems supply oxygen and, thereby, can be used for biosolids aeration before the biosolids are mixed with lime. If storage of unlimed biosolids is contemplated, the designer should check that the air requirement for mixing is sufficient to meet the oxygen demand of the biosolids.

There are disadvantages to both types of mixing systems. Mechanical mixers are subject to fouling with rags, strings, and other debris in the biosolids. Although air spargers may clog, fouling problems are greatly reduced by mixing with air. Ammonia will be stripped from the biosolids with diffused air when mixing is done, producing odor and reducing the fertilizer value of the treated biosolids. However, if nitrogen levels limit land application rates, this stripping of ammonia will reduce land requirements for disposal. A probably minor problem with air mixing is that CO_2 is absorbed by the biosolids/lime mixture, tending to raise the quantity of lime required to reach the desired pH. The selection of the method of mixing should be based on the factors described earlier, coupled with an economic evaluation.

With air mixing, coarse bubble diffusers should be used, mounted along one of the tank walls to induce a spiral-roll mixing pattern. An air supply of 20–30 scfm/1000 ft^3 (20–30 m^3/min/1000 m^3) is required for adequate mixing (43). If the mixing tank is enclosed, ventilation should be sufficient to remove odorous gases stripped from the biosolids during mixing. In many cases, these gases should be treated in an odor control unit before being discharged into the atmosphere.

Mechanical mixer specifications for various tank sizes are presented in Table 8. Sizing is based on two criteria: maintaining the bulk fluid velocity (defined as the turbine agitator pumping capacity divided by the cross-sectional area of the mixing vessel) more than 26 ft/min (8.5 m/min), and using an impeller Reynolds number more than 1000. The tank/mixer combinations in Table 8 are adequate for mixing biosolids with up to 10% dry solids and viscosity of 1000 cp. Impellers on mechanical mixers should be designed to minimize fouling with debris in the biosolids.

5.3. PSRP Treatment to Meet Class B Requirements

PSRP (processes to significantly reduce pathogens) treatment is used to meet class B requirements. The PSRP system (Fig. 1) raises the pH to 12.0 and holds it there for 2 h. Generally screw conveyors are the best to handle biosolids. Modifications to the screw design are required for biosolids, and the actual capacity of the screw conveyor may be as little as 15% of the Conveyor Equipment Manufacturers Association (CEMA) published data. The rpm and conveyor tip speed are important factors for biosolids capacity

Table 8
Mechanical Mixer Specifications for Biosolids Slurries

Tank size (gal)	Tank diameter (ft)	Motor size (hp)	Shaft speed (rpm)	Turbine diameter (ft)
5000	9.5	7.5	125	2.7
		5	84	3.2
		3	56	3.6
15,000	13.7	20	100	3.7
		15	68	4.4
		10	45	5.3
		7.5	37	5.6
30,000	17.2	40	84	4.8
		30	68	5.1
		25	56	5.5
		20	37	6.8
75,000	23.4	100	100	5.2
		75	68	6.2
		60	56	6.6
		50	45	7.3
100,000	25.7	125	84	6.0
		100	68	6.5
		75	45	7.8

Source: US EPA.
Assumptions: Bulk fluid velocity >26 ft/min (8.5 m/min).
Impeller Reynolds number >1000.
Mixing tank configuration:
Liquid depth equals tank diameter.
Baffles with a width of one-twelfth of the tank diameter, placed at 90° spacing.

and conveying. It is recommended that a manufacturer experienced in biosolids conveying be selected to incorporate the specific characteristics of the biosolids, and thus ensure the required capacity and prevent thixotropic conditions. The screw conveyors provide excellent heat transfer and are ideal for addition of supplemental heat, either electrically or from steam/hot water. This can reduce lime usage by an average of 18% (16).

Performance testing should be performed to ensure the entire mass of biosolids is brought to a pH of 12.0. The following criteria should be met: for supplement heating, an insulated mixer should be used; a minimum retention time of 60–90 s thorough mixing of biosolids and lime to a granular consistency; 95% of material discharged, 0.5 in. and smaller.

5.4. PFRP Treatment to Meet Class A Requirements

PFRP (processes to further reduce pathogens) treatment is also used to meet class A requirements. PFRP treatment requires that the biosolids be brought to pH 12.0 and concurrently elevated to 70°C. This has the same configuration as for the PSRP system, but is followed by a pasteurization vessel (Fig. 1). Supplemental heat is required, and insulation of the mixer is very important. Supplemental heat can be provided more economically by electricity than by lime addition as shown in Table 9 (16). Transfer conveyors may be required to move biosolids from the mixer to the pasteurization vessel.

Table 9
Cost of Supplemental Heat by Lime Addition and Electricity[a]

	2004 USD/10,000 BTU/h
Quick lime at 98 USD/t	1
Electricity at 0.065 USD/kW	0.20
Electricity at 0.13 at 13 USD/kW	0.40

[a] Cost expressed in year 2004 USD = 1990 costs × 1.31 (*see* Appendix).

These should be insulated to retain heat. No heat addition is required unless the transport distance is 50 ft or longer, and then only a minor amount that may be supplied by heat tracing. The pasteurization vessel should provide a retention time of 30–60 min, with a minimum of 30 min at 70°C or more. The biosolids-lime temperature continues to rise for 15–30 min after discharge from the biosolids–lime mixer. The pasteurization vessel must be insulated and heart traced as needed to achieve a minimum of 30 min residence time at 70°C. The flow pattern through the vessel should prevent short-circuiting and ensure the 30 min minimum residence time. Generally the vessel should be round and tapered to a conical opening of 24 in. wide and 6 ft long.

Ideally, the system should be designed to operate in either the PSRP or PFRP modes. This will give flexibility to operate in the PSRP mode when producing for landfill or in PFRP mode when producing for public distribution (16).

6. COST AND ENERGY USAGE

This section discusses cost and energy usage for lime stabilization systems.

6.1. Capital and Operating Costs

The following items must be considered in estimating costs for any lime stabilization facility (12):

a. Processing equipment purchase and installation.
b. Product curing and storage facilities.
c. Loading facilities.
d. Transport of product to point of use.
e. Royalty and operating fees for proprietary processes.
f. Equipment maintenance and fuel.
g. Alkaline additive.
h. Labor.
i. Odor control equipment and chemicals.
j. Marketing costs/revenues.
k. Regulatory compliance, such as permit applications, site monitoring, biosolids analyses, and regulatory record keeping and reporting.

Lime stabilization systems are relatively uncomplicated facilities operated with the skills found in typical wastewater treatment plant personnel. Labor requirements include heavy equipment operators, maintenance personnel, and instrumentation/computer operators (12). The caustic nature of the alkaline additive requires higher maintenance on these systems than on stabilization systems that do not involve caustic

Table 10
Annual Costs for Lime Stabilization Facilities Expressed in 2004 USD[a]

	Treatment plant capacity (MGD)		
Item	1	4	40
Capital	22,600	64,700	187,500
Operation and maintenance	27,100	77,200	553,400
Total	49,700	141,900	740,900
Unit cost (USD/t dry biosolids)	116.50	84.50	44.10

Source: US EPA.
[a] Cost expressed in year 2004 USD = 1978 costs × 2.15 (*see* Appendix).

materials. Proper design and operation of the mixing equipment is necessary to ensure a consistent, homogeneous product (12).

It is difficult to estimate the cost of stabilizing biosolids with alkaline materials without specific details, such as wastewater solids characteristics and quantities. One study, in 1996, estimated cost for class A alkaline stabilization ranging from 139 to 312 USD/dry t of wastewater solids processed, by facilities designed to serve wastewater treatment plants ranging in capacity from 10 to 60 MGD. This estimated range demonstrates the economy of scale associated with larger systems. The capital costs cited in this same study ranged from 1.5 to 4 million USD and annual costs were estimated to range from 1 to 4 million USD. This study concluded that alkaline stabilization is less expensive than composting or thermal drying (45).

The incremental capital cost for meeting class A requirements through alkaline stabilization is the lowest among stabilization alternatives such as thermal drying and anaerobic and aerobic digestion. The incremental unit cost (including capital and operation and maintenance) associated with creating a class A product from a system currently making a class B product and serving a 5 MGD wastewater treatment facility was estimated to be 39 USD/dry t in 1998. Again, this is significantly less than the unit cost to increase pathogen treatment through aerobic or anaerobic digestion, which were cited as 88 and 103 USD/t, respectively (22). Some generators of alkaline-stabilized biosolids sold the product in year 2000 for about 3–5 USD/wet t (12).

Cost estimates for the construction and operation of three different size lime stabilization systems are summarized in Table 10. A comparison of these costs shows that there is a large economy of scale, especially for the capital costs (18). Operation and maintenance expenses, particularly those for lime, are more closely related to the quantity of biosolids treated.

Comparisons of the cost of lime treatment with other stabilization methods must take into account that the addition of lime increases the quantity of solids to be handled after stabilization. In contrast, biosolids actually decrease during anaerobic and aerobic digestion. This difference between stabilization methods can have an important effect on costs for final disposal of biosolids. The magnitude of this cost difference is site-specific and depends on such factors as the method of disposal and the distance to the disposal site.

6.2. Energy Usage

Energy is required during both the construction and operation of a lime stabilization system. During operation of a lime stabilization facility, the principal direct use of energy is electricity for mixing the lime/biosolids mixture. A rough estimate of the annual energy requirement for mixing with diffused air is 290 kWh/yr/cfm of blower capacity (based on continuous duty). This estimate was made assuming a 6 psig (0.4 kg/m^2) pressure boost, standard inlet conditions, and an overall compressor/motor efficiency of 60%. About 1 hp of mechanical mixing requires about 6500 kWh of electricity/yr. These mixing energy demands can be expressed in terms of a primary fuel requirement (that is, fuel oil, coal, and natural gas) by applying a conversion factor of 10,700 Btu (2700 kg cal) per kWh of electricity. This factor assumes a fuel conversion efficiency of 35% at the power plant and a transmission efficiency of 91% (18).

A large amount of energy is used in the production of quicklime. Quicklime (CaO) is produced by burning limestone ($CaCO_3$) in kilns. This process, termed calcination, is illustrated in the following reaction:

$$CaCO_3 \xrightarrow{\text{heat}} CaO + CO_2 \tag{4}$$

The national average energy consumption for all quicklime production is about 7.0 million Btu/t of quicklime (1.9 × 10^6 kg cal/T) (46). This figure is decreasing because modern plants, using large and more efficient kilns, are be able to produce 1 t of quicklime with about 5.5 million Btu (1.5 × 10^6 kg cal/T). Here T = 1000 kg.

6.3. Design Comparison for Lime-Only and Supplemental Heating Pasteurization

In 1993 a pilot test compared lime-only and supplemental heating pasteurization (47). The objective was to produce lime stabilized biosolids for spreading on farm land. The biosolids were to comply with class A alternative 1 pathogen reduction (503 regulations, 30 min at 70°C). The lime-only pasteurization method used a biosolids/lime mixer and insulated pasteurization vessel provided by the F. B. Leopold Company, Inc. The alternative system (RDP En-Vessel Pasteurization) used an electrically heated biosolids/lime mixer and an electrically heated pasteurization vessel from the RDP Company. The cost and performance basis used the following criteria:

Electrical cost	0.055 USD/kWh (in 1993)
Lime cost	75 USD/t (in 1993)
Motor efficiency	90%
Biosolids production	2600 dry lb/h, 40 h/wk, 52 wk/yr

It was determined that the lime-only process should use a lime feed rate of 106%; and the supplemental heating method, 25.3%, plus 0.067 kWh of added heat/lb of biosolids. Table 11 compares capital and operating costs for the two cases.

Other factors assessed included the appearance, fertilizer and lime content, and the land area disposal requirement. The lime-only product had a lighter, more appealing appearance. The lime-only product had higher lime content, and the supplemental

Table 11
Capital and Operating Costs for Lime Only and Supplemental Heating Pasteurization[a]

Cost	Lime-only pasteurization 2004 USD	Supplemental heating pasteurization 2004 USD
Lime	264,410	63,110
Electrical (hp)	4490	7000
Electrical-heating	NA	24,510
Total annual operating cost	268,900	94,620
Present worth at 6–5/8% over 10 yr	1,921,840	676,230
Capital cost	848,700	1,270,440
Total cost	2,770,540	1,946,670

[a] Cost expressed in year 2004 USD = 1993 costs × 1.23 (*see* Appendix).

Table 12
Cost Comparison Summary for Lime Only and Supplemental Heating Pasteurization[a,b]

Parameter	Supplemental heating pasteurization cost 2004 USD	Lime only pasteurization cost 2004 USD
Lime cost per dry ton biosolids	23.33	97.79
Power cost per dry ton biosolids	11.65	1.66
Total cost	34.98	99.45
Land required per dry ton biosolids	0.59	0.96

[a] Cost expressed in year 2004 USD = 1993 costs × 1.23 (*see* Appendix).
[b] RDP En-Vessel Pasteurization™.

heating product had more nitrogen. Estimation of the disposal requirement assumed a nitrogen utilization rate of 130 lb/acre/yr and lime loading rate of 2000 lb/acre/yr. Table 12 summarizes costs and land disposal requirements (47).

7. DESIGN EXAMPLE

This example illustrates the layout and sizing of the major components in a lime stabilization system. For this example, it is assumed that the treatment plant has a capacity of approx 8 MGD (350 L/s) and provides secondary treatment to typical municipal wastewater. A mixture of primary biosolids and thickened waste-activated sludge is to be stabilized with lime, then mechanically dewatered, and ultimately spread onto land.

7.1. Design Loading

Biosolids (sludge) production estimates for two flow conditions, average and peak day, are listed in Table 13. The peak loading is listed because critical components must

Table 13
Design Loading Assumptions

	Flow condition	
Parameter	Average flow	Peak daily flow
Sludge production (lb dry solids/d)		
Primary sludge	10,000	15,000
Waste-activated sludge	5000	7500
Solids concentration (%)		
Primary sludge	5	4
Waste-activated sludge	4	3.5
Sludge volume (ft^3/d)		
Primary sludge	3200	6010
Waste-activated sludge	2000	3430

be sized to meet this critical condition. Biosolids concentrations and the resulting biosolids volumes are also included in Table 13.

7.2. System Description

The conceptual design for the lime stabilization system is presented in Fig. 6. Before stabilization, all biosolids are passed through an in-line grinder. This conditioning improves biosolids mixing and flow characteristics, protects downstream pumping and dewatering equipment, and eliminates unsightly conditions (such as rags, sticks, and plastic) at the disposal site. Two batch-mixing tanks are provided, each with the capacity to treat the total biosolids produced in an 8 h shift during peak day conditions. While one tank is filling, biosolids in the other are treated with lime, mixed for 30 min, and then discharged to the dewatering equipment. Because the mixing tanks are sized for peak conditions, they can provide some short-term storage for treated biosolids during periods of lower loading. Design of an actual facility should take into consideration the operating schedule for dewatering and disposal.

In this example, it is assumed that dewatering is operated continuously and therefore only minimal inline storage is required. However, if dewatering equipment was operated for two shifts, and serviced during the third, at least 8 h of storage would be required. Air discharged through coarse bubble diffusers is used to mix the biosolids with the lime slurry. Air mixing is started as raw biosolids are first added to the tank to keep the biosolids from turning septic and producing odor. When the tank is filled, lime is added and mixing is continued for at least 30 min.

To reduce odor, the mixing tanks are covered, and gases stripped from the biosolids during mixing are removed in an odor control unit. This unit is a packed bed scrubber. The scrubbing solution is dilute sulfuric acid. Ammonia gas is absorbed by the sulfuric acid solution. All wetted parts are constructed of acid-resistant materials.

Quicklime is used in this installation. A bulk storage silo, with capacity to hold a 30 d lime requirement under average conditions, supplies lime to two volumetric feeders. Each feeder measures out quicklime to a shaker, where the lime is hydrated, slurried, and discharged into the mixing tank. The lime dose is sufficient to maintain the biosolids more than pH 12.5 for 30 min.

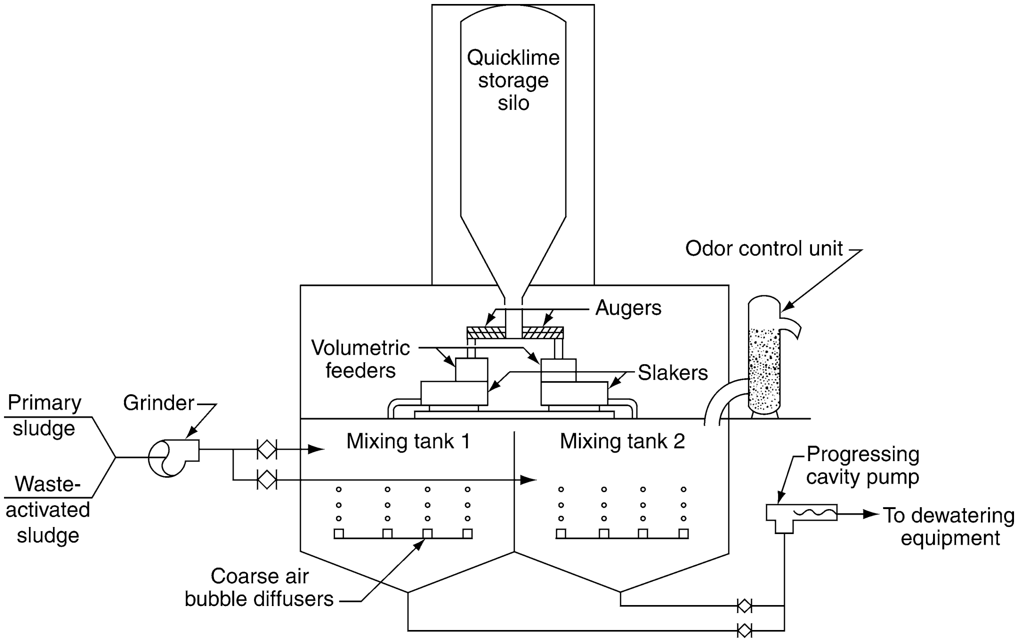

Fig. 6. Design for a lime stabilization facility (*Source:* US EPA).

7.3. Component Sizing

7.3.1. Mixing Tank

Volume requirement (V): peak-day biosolids production is shown in Table 13.

$$
\begin{aligned}
V &= [(8\ \text{h/tank})/(24\ \text{h/d})]\ (6010\ \text{ft}^3/\text{d} + 3430\ \text{ft}^3/\text{d})\\
&= 3150\ \text{ft}^3/\text{tank}\\
&= 89\ \text{m}^3/\text{tank}.
\end{aligned}
$$

Tank surface area (A): (assume 10 ft liquid depth)

$$A = 3150\ \text{ft}^3/10\ \text{ft} = 315\ \text{ft}^2 = 39.3\ \text{m}^2.$$

Tank dimensions: (assume 2 ft freeboard)

$18 \times 18 \times 12\ \text{ft}^3$
$(5.4 \times 5.4 \times 3.7\ \text{m}^3)$.

7.3.2. Air Mixing System

Sizing criterion: 30 cfm/1000 ft^3.
Blower capacity (Q): (one blower per tank)

$$
\begin{aligned}
Q &= (3150\ \text{ft}^3)\ (30\ \text{cfm/l000 ft}^3)\ \text{tank}\\
&= 95\ \text{cfm/blower} = 2.6\ \text{m}^3/\text{min/blower}
\end{aligned}
$$

where cfm is ft^3/min.

7.3.3. Lime Storage

Sizing criterion: 30 d storage during average loading.

a. Quicklime characteristics:
   Purity: CaO (90%).
   Bulk density: 55 lb/ft^3.

b. Lime dosage:
   Primary biosolids: 0.12 lb $Ca(OH)_2$/lb dry solids.
   Activated sludge: 0.30 lb $Ca(OH)_2$/lb dry solids.

c. Average daily lime requirement (W):
   Expressed as hydrated lime:

$$\begin{aligned} W_{Ca(OH)_2} &= (10{,}000 \text{ lb/d})\,(0.12 \text{ lb/lb}) + (5000 \text{ lb/d})\,(0.30 \text{ lb/lb}) \\ &= 2700 \text{ lb } Ca(OH)_2\text{/d} \\ &= 1230 \text{ kg/d} \end{aligned}$$

   Expressed as purchased quicklime (90% purity):

$$\begin{aligned} W_{CaO} &= 2700 \text{ lb } Ca(OH)_2\ [(56 \text{ lb CaO/mole})/(74 \text{ lb } Ca(OH)_2\text{/mole})]\ (100/90) \\ &= 2270 \text{ lb CaO/d} \\ &= 1030 \text{ kg/d} \end{aligned}$$

   Storage requirement (V_s):

$$\begin{aligned} V_s &= [(2270 \text{ lb/d})/(55 \text{ lb/ft}^3)]\ (30 \text{ d}) \\ &= 1240 \text{ ft}^3 \\ &= 35 \text{ m}^3 \end{aligned}$$

7.3.4. Slaker

Sizing criterion: Ability to dose one batch in 15 min.
Slaker capacity (C):

$$\begin{aligned} C &= [(2270 \text{ lb CaO/d})/(3 \text{ batches/d})]\ (1 \text{ batch/15 min}) \\ &= 50 \text{ lb CaO/min} \\ &= 23 \text{ kg/min} \end{aligned}$$

NOMENCLATURE

A	Tank surface area (m^2 [ft^2])
C	Slaker capacity (kg CaO/min [lb/min])
Q	Blower capacity, (m^3/min [cfm]).
t	English ton; 1 t = 2000 lb
T	Metric tonne; 1 T = 1000 kg
V	Tank volume (m^3 [ft^3])
V_s	Storage requirement (m^3 [ft^3])
W	Lime requirement (kg/d [lb/d])
W_{CaO}	CaO requirement (kg/d [lb/d])
$W_{Ca(OH)_2}$	$Ca(OH)_2$ requirement (kg/d [lb/d])

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APPENDIX

United States Yearly Average Cost Index for Utilities[a]

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

[a] Extracted from ref. 44.

8
Pressurized Ozonation

Lawrence K. Wang and Nazih K. Shammas

CONTENTS

1. INTRODUCTION

Increasing populations and improving standards of living are placing increasing burdens on water resources. The preservation of our limited natural water supplies and, in the not too distant future, the necessity for direct recycling of water in some parts of the country will demand improved technology for the removal of contaminants from wastewater. The contaminants in wastewater are many and continually varying, and they are not well-characterized according to chemical species. Commonly the level of organic contamination is expressed by biochemical oxygen demand (BOD), chemical oxygen demand (COD), or total organic carbon (TOC). Ozone and oxygen are powerful oxidants, which can oxidize many contaminants in wastewater and sludge biosolids. Ozone is more powerful than oxygen, but it is an unstable material, which must be generated at the point of use. Ozone has been used for disinfecting drinking water in European countries for many years. It has also been used for treating some special industrial wastes, notably for removing cyanides and phenols. Since 1980, ozone started to be used for wastewater, industrial wastes, and sludge treatment on a large scale (1–6). Oxidative purification and disinfection with ozone as a tertiary wastewater treatment or sludge treatment has a number of inherent advantages which are as follows:

From: *Handbook of Environmental Engineering, Volume 6: Biosolids Treatment Processes*
Edited by: L. K. Wang et al. © The Humana Press Inc., Totowa, NJ

a. Reduction in BOD and COD.
b. Reduction of odor, color, turbidity, and surfactants.
c. Pathogenic organisms are destroyed.
d. The treatment products are beneficial.
e. The effluent water has a high dissolved oxygen (DO) concentration.

The relative high cost of ozone generation requires a high ozone-utilization efficiency if ozone treatment is to be economically competitive. A principal disadvantage to the use of ozone in waste treatment is its cost. However, recent advances in ozone generation have rendered the ozonation process more competitive.

This chapter deals with two newly developed oxygenation–ozonation (Oxyozosynthesis™) systems for wastewater and sludge treatment. Each treatment scheme consists of a wet well for flow equalization and pH adjustment, a hyperbaric reactor for oxygenation and ozonation, a flotation clarifier for degasification and solid–water separation, and a filter belt press for final sludge-dewatering. Special emphasis is placed on theory, kinetics, and disinfection effect of ozonation and oxygenation (7–12).

1.1. Oxyozosynthesis Sludge Management System

As shown in Fig. 1 and Fig. 2, the new sludge management system consists of the following unit operations and processes: sludge production from clarifiers, flow equalization, and pH adjustment in a wet well, oxygenation–ozonation in a hyperbaric reactor vessel (Fig. 3), flotation, dewatering in a belt press, and resource recovery of final product as fuel or for land application.

A full-scale Oxyozosynthesis sludge management system was installed at the West New York Sewage Treatment Plant (WNYSTP), West New York, NJ. The plant treats domestic wastewater flow of 10 MGD and produces 22,000 gpd of primary sludge. Primary raw sludge is pumped from sumps located at the bottom of the primary sedimentation clarifiers by means of two positive displacement pumps to a sludge grinder, then to the wet well. As the wet well is being filled with grinded sludge, a 10% sulfuric acid solution is added to adjust the pH value to a 3.5–4.0 by a chemical metering pump. A mechanical mixer and a pH meter are mounted in the wet well for proper mixing and pH monitoring, respectively. Following acidification, the sludge is pumped by a progressive cavity pump to one of the two batch-operated hyperbaric reactor vessels, each capable of treating 1500 gal of sludge in 90 min by oxygenation and ozonation. To start up each reactor vessel, the pressure in the reactor is first built up to 40 psig with liquid oxygen and then up to 60 psig with ozone. There are two operational modes:

a. **Continuous oxygenation–ozonation.** After the startup with oxygen and ozone, ozone is continuously fed into the reactor for a total of 90 min. The pressure is maintained at 60 psig by bleeding off (or recycling) the excess gas.
b. **Noncontinuous oxygenation–ozonation.** After the startup with oxygen and ozone, ozone is then shut off to let the reactor be isolated and maintained as such for 90 min.

During the 90 min contact time in the oxygenation–ozonation reactor, pathogenic bacteria, viruses, total suspended solids, and volatile suspended solids in the sludge are all significantly reduced. The reactor effluent is then released (at a flow rate of approx 1500 gal/90 min) into an open flotation unit where DO, ozone, and carbon dioxide gases

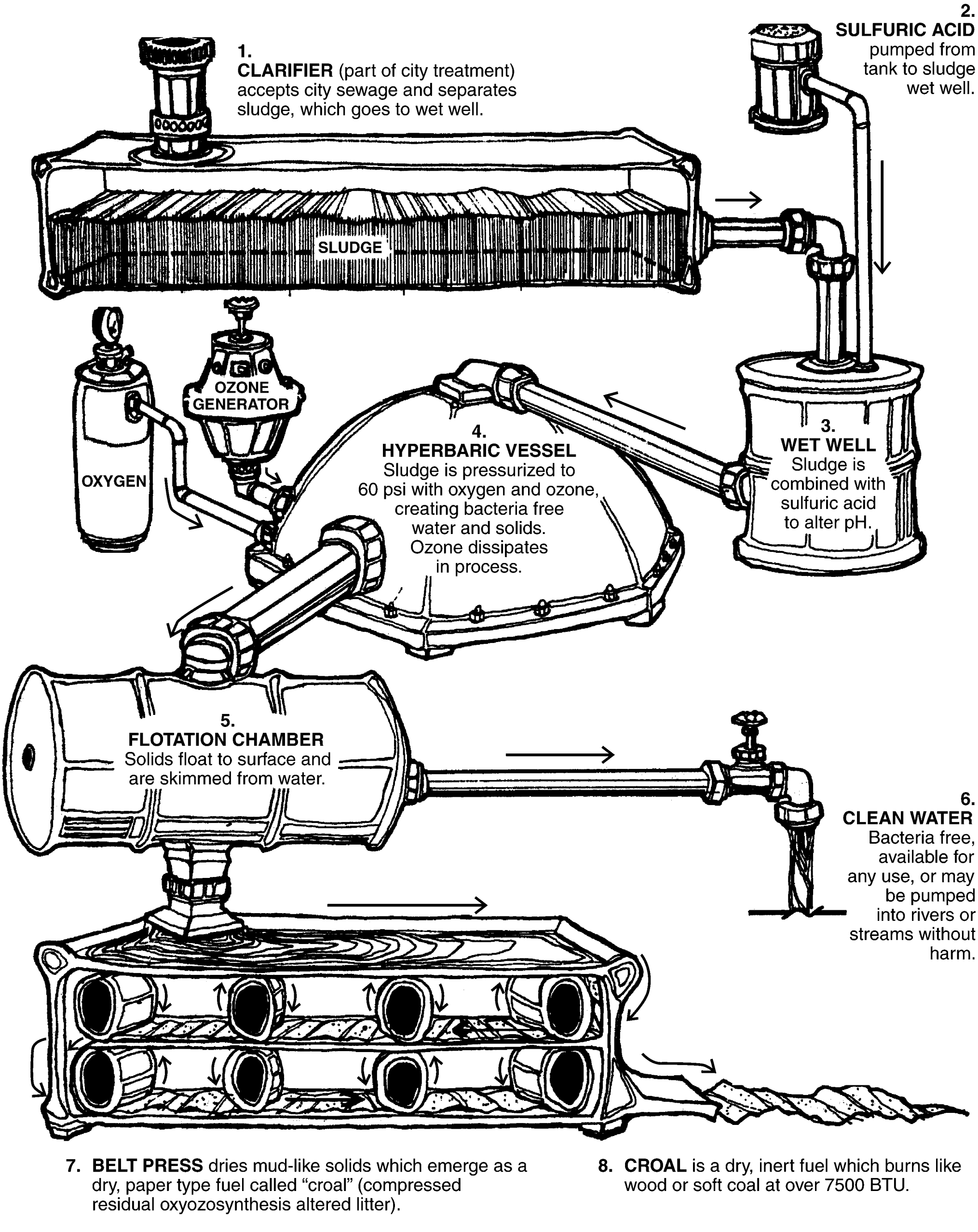


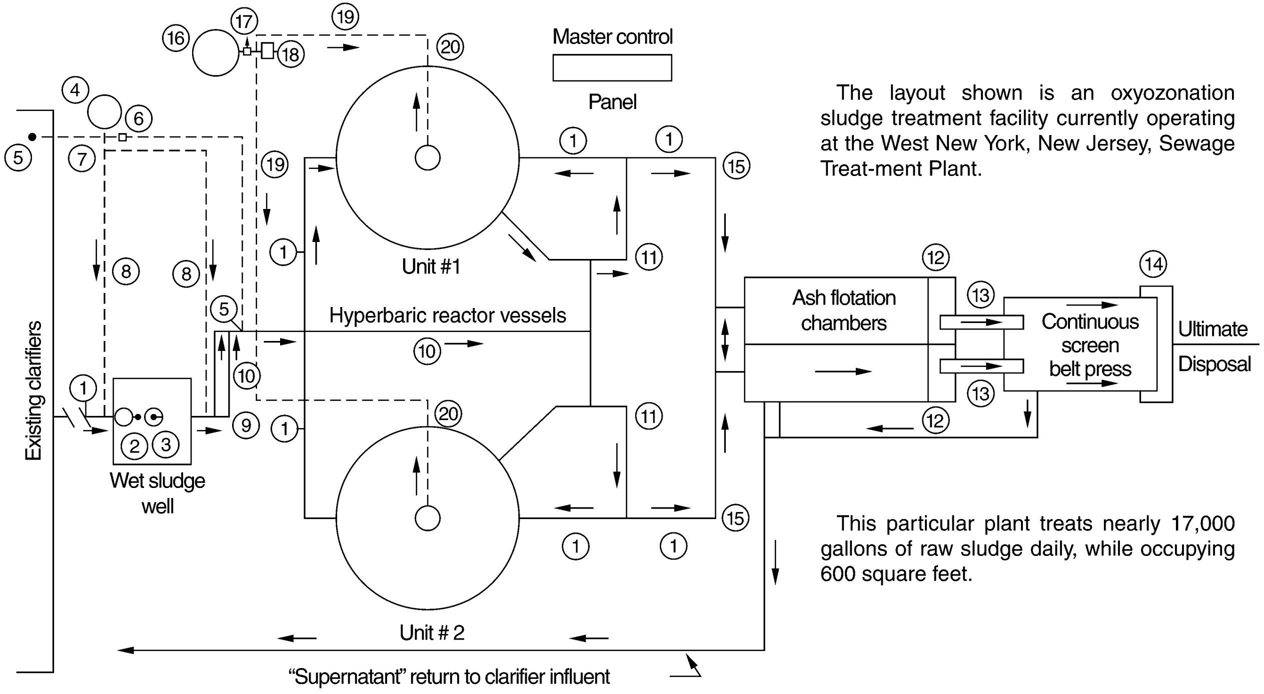
Fig. 1. General view of oxygenation–ozonation (Oxyozosynthesis) System.

come out of solution and form tiny bubbles, which adhere to the residual suspended solids causing them to float and get thickened at the top of the unit. The flotation unit is equipped with revolving paddles (or scoops) that transport these floating solids onto a subsequent filter belt press for sludge dewatering. The subnatant liquor is recycled to the head of the sewage treatment plant for further treatment with the incoming wastewater flow.

WASTEWATER SLUDGE DISPOSAL SYSTEM

LEGEND

1. Pinch type flow control valve
2. Sludge grinder
3. Mixer
4. Chemical solution storage tank
5. PH probe
6. PH control device
7. Chemical solution feed pump
8. Chemical feed lines
9. Variable speed progressive cavity pump
10. Influent pump
11. Progressive cavity recirculation pump
12. Floated sludge removal device
13. Sludge conveyor
14. Sludge disposal bin (Portable)
15. Auxillary sludge removal
16. Oxygen storage tank
17. Oxygen supply control device
18. Ozone generator
19. O_2 and O_3 Feed line
20. Gas recirculation and Gas feed line



The layout shown is an oxyozonation sludge treatment facility currently operating at the West New York, New Jersey, Sewage Treat-ment Plant.

This particular plant treats nearly 17,000 gallons of raw sludge daily, while occupying 600 square feet.

Fig. 2. Flow diagram of oxyozosynthesis sludge management system.

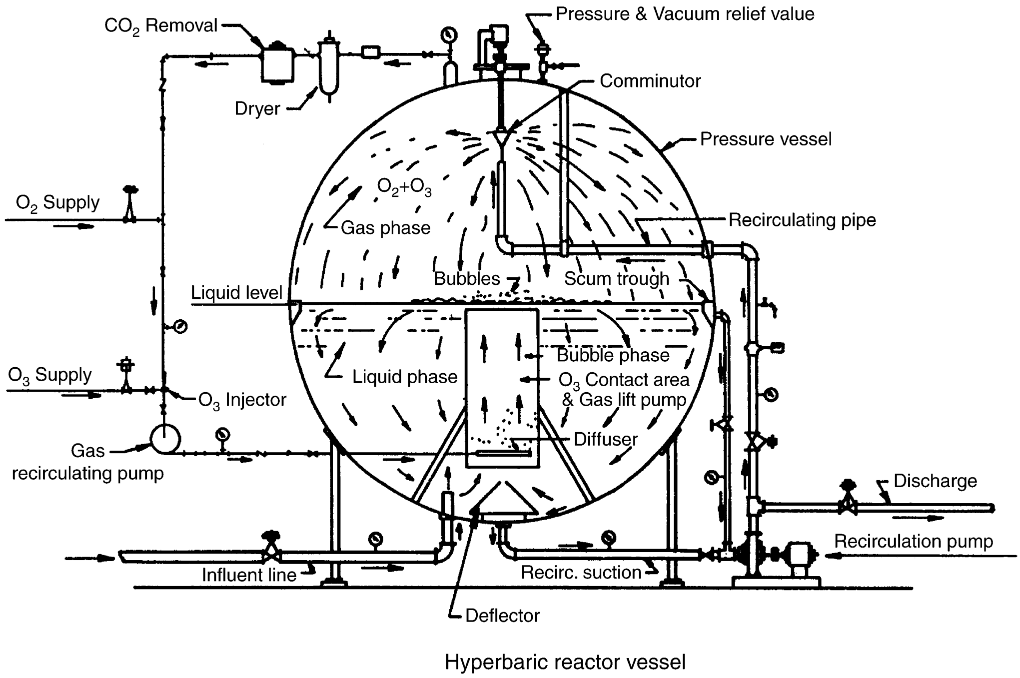


Fig. 3. The hyperbaric reactor vessel.

The filter belt press produces a dry high-nutrient sludge cake with low metal content and high Btu value. The sludge cake can be recycled by spreading on agricultural land, reused as a fuel source or disposed off in landfill. The dry sludge can also be reused as secondary fiber in paper manufacturing, or as raw material for building blocks.

1.2. Oxyozosynthesis Wastewater Reclamation System

As shown in Fig. 4, the new wastewater reclamation system consists of the following unit operations and processes: wastewater collection and preliminary treatment (bar screens and grit chambers), flow equalization and pH adjustment in a wet well, oxygenation–ozonation in a hyperbaric reactor vessel, dissolved gas flotation (DGF), and filtration.

A pilot-scale Oxyozosynthesis wastewater reclamation system was installed at the Lenox Institute of Water Technology, Lenox, Massachusetts. The pilot plant treats a wastewater flow of 6 gpm and produces small amount of sludge. Raw wastewater is pumped from sumps located at the bottom of the grit chambers by means of positive displacement pumps to a wet well. As the wet well is being filled with the raw wastewater, a 10% sulfuric acid solution is added to adjust the pH value to a 3.5–4.0 by a chemical metering pump. A mechanical mixer and a pH meter are mounted in the wet well for proper mixing and pH monitoring, respectively.

From the wet well, a progressive cavity pump delivers the acidified wastewater to a batch-operated hyperbaric reactor vessel capable of treating 100 gal of wastewater in 30–60 min depending on the characteristics of the wastewater. To start up the reactor vessel, the pressure in the reactor is built up to 40 psig with liquid oxygen first, and then up to 60 psig with ozone. There are two operational modes:

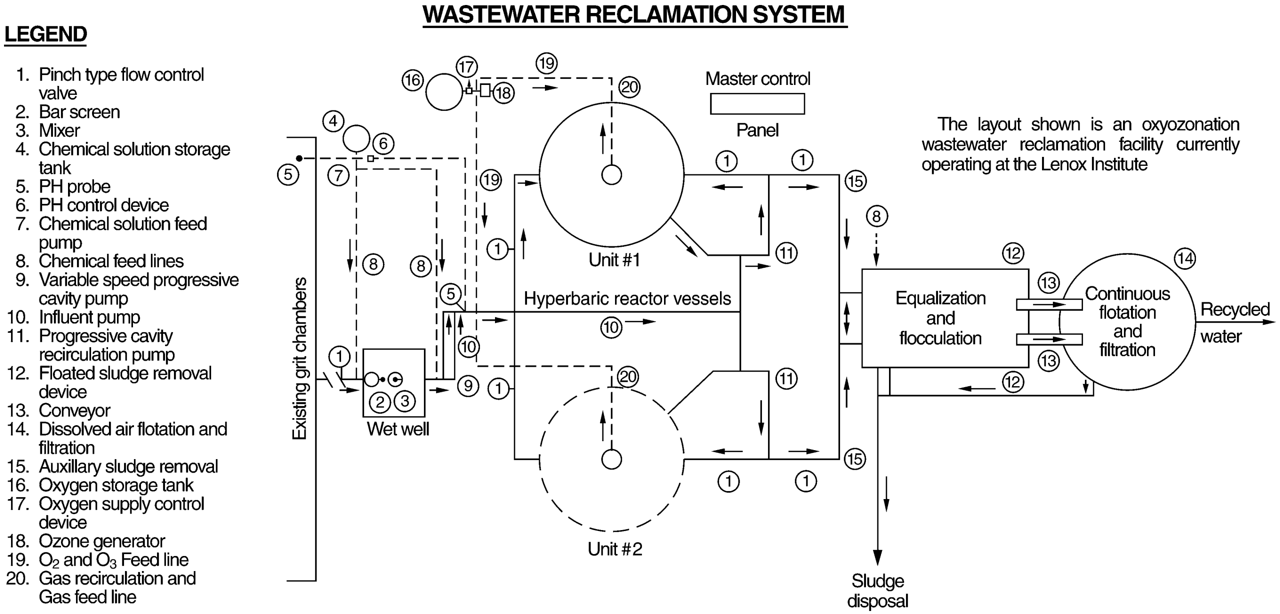


Fig. 4. Flow diagram of oxyozosynthesis wastewater reclamation system.

a. **Continuous oxygenation–ozonation.** After the startup with oxygen and ozone, ozone is continuously fed into the reactor for a total of 30–60 min. The pressure is maintained at 60 psig by bleeding off (or recycling) the excess gas.
b. **Noncontinuous oxygenation–ozonation.** After the startup with oxygen and ozone, ozone is then shut off to let the reactor be isolated and maintained as such for 30–60 min.

During the 30–60 min contact time in the oxygenation–ozonation reactor, pathogenic bacteria, viruses, total suspended and volatile suspended solids, phenols, cyanides, manganese, and so on, in wastewater are all significantly reduced. The reactor effluent is released into a DGF unit in which flocculant(s) can be added and the dissolved gases come out of aqueous phase forming tiny bubbles, which adhere to the flocs and residual suspended solids causing them to float to the top of the unit. Heavy metals, iron, phosphate, humic acids, hardness, toxic volatile organics, and so on, will react with the flocculant(s) to form insoluble flocs that are floated. The flotation unit is equipped with revolving paddles (or scoops) that transport these floating solids, onto a subsequent filter belt press for final sludge dewatering. A dual-media filter further polishes the subnatant clarified water.

The filter effluent quality is not far from that of potable water, having extremely low color, turbidity, suspended solids, hardness, iron, manganese, trihalomethane precursor (humic acid), heavy metal, volatile organics, phenol, cyanide, and so on. The product water is suitable for reuse for industrial and agricultural purposes. Further treatment of the final filter effluent by adsorption on activated carbon is optional.

2. DESCRIPTION OF PROCESSES

2.1. Ozonation and Oxygenation Process

Ozone gas is sparingly soluble in water. The solubility of ozone in water increases with its increasing partial pressure, decreasing water pH, and decreasing temperature. However, oxidation rate increases with increasing temperature. For economic operation of the hyperbaric oxygenation–ozonation reactor, it is operated at room temperature and a pressure in the range of 40–60 psig, the influent liquid sludge pH value is reduced with sulfuric acid down to a range of 3.5–4.0.

The addition of oxygen at 40 psig and ozone at 60 psig ensure proper partial pressure for solubilizing both oxygen and ozone gases in the sludge. Both DO and ozone act to chemically oxidize reducing pollutants found in the liquid sludge thus reducing BOD and COD, which result in the formation of oxygenated organic intermediates and end products. Ozonation–oxygenation treatment also reduces color and odor in waste sludge. Because there is a wide range of ozone reactivity with the diverse organic content of wastewater, both the required ozone dose and reaction time are dependent on the quality of the influent to the ozonation process. Generally, higher doses and longer contact times are required for ozone oxidation reactions than are required for wastewater disinfection using ozone. Ozone tertiary treatment might eliminate the need for a final disinfection step. Ozone breaks down to elemental oxygen in a relatively short period of time (its half-life is about 20 min). Consequently, it must be generated on site using either air or oxygen as the feed gas. Ozone generation utilizes a silent electric arc or corona through which air or oxygen passes, and yields ozone in air/oxygen mixture, the percentage of

Table 1
Effectiveness of Ozone as an Oxidant

	COD (mg/L)		BOD_5 (mg/L)		TOC (mg/L)	
Ozone Dosage (mg/L)	Influent	Effluent	Influent	Effluent	Influent	Effluent
50	318	262	142	110	93	80
100	318	245	142	100	93	77
200	318	200	142	95	93	80
325	318	159	142	60	93	50
50	45	27	13	7	20.5	15.5
100	45	11	13	3	20.5	9
200	45	5.5	13	1.5	20.5	5

Source: US EPA.

ozone being a function of voltage, frequency, gas flow rate, and moisture. Automatic devices are commonly applied to control and adjust the ozone generation rate.

For sludge treatment or wastewater reclamation, it is a developing technology. Recent developments and cost reduction in ozone generation and ozone dissolution technology make the process very competitive. A full-scale application is currently in the demonstration stage at the WNYSTP, West New York, NJ. If oxygen-activated sludge is employed in the system, ozone treatment may be even more economically attractive, because a source of pure oxygen is available facilitating ozone production.

For poor quality wastewater or sludge with extremely high COD, BOD, and/or TOC contents (more than 300 mg/L), ozone treatment can be economical only if there is adequate pretreatment. The process will not produce any halogenated hydrocarbons. Table 1 shows the reduction of overall COD, BOD, and TOC, achieved in the United States Environmental Protection Agency (US EPA) controlled tests after a 90 min contact time with ozone oxidation. Beyond the 70% COD removal level, the oxidation rate is significantly slowed. In laboratory tests, COD removal never reaches 100% even at a high ozone dose of 300 mg/L.

As a disinfectant with common dosages of 3–10 mg/L, ozone is an effective agent for deactivating common forms of bacteria, bacterial spores, and vegetative microorganisms found in wastewater, as well as eliminating harmful viruses. Additionally, ozone acts to chemically oxidize materials found in the wastewater and sludge, forming oxygenated organic intermediates and end products. Further, ozone treatment reduces wastewater color and odor. Ozone disinfection is applicable in cases where chlorine disinfection might produce potentially harmful chlorinated organic compounds. If oxygen-activated sludge is employed in the system, ozone disinfection is economically attractive, because a source of pure oxygen is available facilitating ozone production. However, ozone disinfection, does not form a residual that will persist and can be easily measured to assure adequate dosage. Ozonation may not be economically competitive with chlorination under nonrestrictive local conditions.

Easily oxidizable wastewater organic materials consume ozone at a faster rate than disinfection; therefore, effectiveness of disinfection is inversely correlated with effluent quality but directly proportional to ozone dosage. When sufficient concentration is

Table 2
Effectiveness of Ozone as a Disinfectant

Source	Influent	Dose (mg/L)	Contact time (min)	Effluent residual
US EPA	Secondary effluent	5.5–6	≤1	<2 Fecal coliforms/100 mL
US EPA	Secondary effluent	10	3	99% Inactivation of fecal coliform
US EPA	Secondary effluent	1.75–3.5	13.5	<200 Fecal coliforms/100 mL
US EPA	Drinking water	4	8	Sterilization of virus
WNYSTP	Primary sludge	NA	60	More than 99% inactivation of fecal coliform
SIT/LI	Secondary sludge	NA	60	More than 99% inactivation of fecal coliform

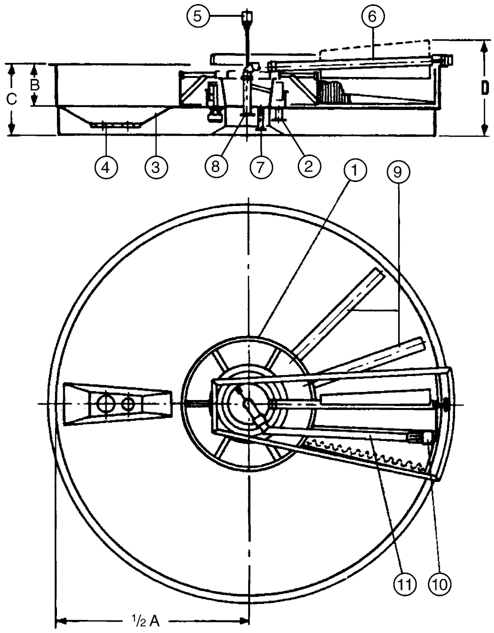
Source: US EPA.

introduced, ozone is a more complete disinfectant than chlorine. Results of disinfection by ozonation have been reported by various sources, which are summarized in Table 2.

2.2. Flotation Process

DGF is mainly used to remove suspended and colloidal solids by flotation resulting from the decrease in their apparent density. The influent feed liquid can be raw water, wastewater, or liquid sludge.

The flotation system consists of four major components: gas supply, pressurizing pump, retention tank, and flotation chamber. According to Henry's Law, the solubility of gas in aqueous solution increases with increasing pressure. A pressurizing pump is used to saturate the feed stream with gas at pressures, several times the atmospheric pressure (25–70 psig). The pressurized feed stream is held at this high pressure for about 0.5–3 min in a retention tank (hyperbaric vessel) designed to provide the required time for dissolution of gas into the treatment stream. Following the retention vessel, the stream is released back to atmospheric pressure in the flotation chamber. Most of the pressure drop occurs downstream from a pressure-reducing valve and in the transfer line between the retention vessel and the flotation chamber so that the turbulent effect of depressurization is minimized. The sudden reduction in pressure in the flotation chamber results in the release of microscopic gas bubbles (average diameter 80 μm or smaller) that attach themselves to the suspended and colloidal particles present in water. This results in an agglomeration, which because of its entrained gas gives a net combined specific gravity less than that of water whose consequence is the flotation phenomenon. The vertical rising rate of gas bubbles ranges between 0.5 and 2 ft/min. The floated materials rise to the surface of the flotation chamber where they are continuously scooped by specially designed flight scrapers or other skimming devices. The surface sludge layer or float can in certain cases attain a thickness of several inches and be relatively stable. The layer thickens with time, but undue long delays in removal will cause release of particulates back to the liquid. The clarified effluent is usually drawn off from the bottom of the flotation chamber, which can be recovered for reuse or for final disposal. Figures 5 and 6 illustrate up-to-date DGF systems using single cell and double cell, respectively. The flotation system is known as dissolved air flotation (DAF) if only



1. Rotating center section
2. Clarified water outlet
3. Settled sludge sump
4. Settled sludge outlet
5. Rotary contact
6. Spiral scoop
7. Floated sludge outlet
8. Unclarified water inlet
9. Clarified water extraction pipes
10. Gear motor
11. Distribution duct

A Diameter of supracell
B Depth of supracell tank
C Depth of supracell tank with bottom support
D Minimum overall height of supracell

TYPE		DIMENSIONS						FLOW		
A ft	A mm	B in	B mm	C in	C mm	D in	D mm	m^3/min	US GPM	m^3/h
8	2400	23.5	600	33	850	45	1150	0,56	148	34
10	3200	23.5	600	33	850	49	1250	1,00	263	60
12	3900	25.5	650	35	900	51	1300	1,50	394	90
15	4500	25.5	650	37	950	57	1450	2,00	525	120
18	5500	25.5	650	37	950	58	1480	3,00	789	180
20	6100	25.5	650	37	950	61	1560	3,65	961	219
22	6700	25.5	650	37	950	62	1580	4,40	1160	264
24	7200	25.5	650	37	950	63	1600	5,08	1340	305
27	8100	25.5	650	37	950	67	1700	6,44	1695	386
30	9000	25.5	650	37	950	71	1820	7,95	2090	477
33	10000	25.5	650	37	950	72	1840	9,80	2580	588
36	11000	25.5	650	37	950	73	1860	11,87	3125	712
40	12200	26	660	38	960	76	1920	14,60	3840	876
44	13400	27	685	39	985	78	1980	17,60	4630	1056
49	14800	27	685	39	985	82	2070	21,50	5650	1290
55	16800	27	685	39	985	87	2200	27,70	7290	1662

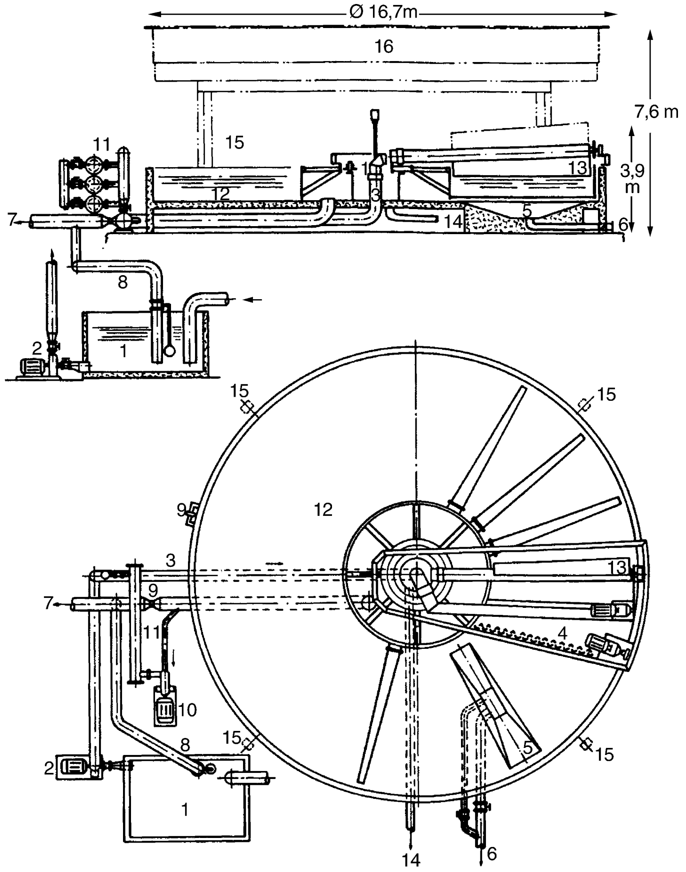
Fig. 5. A single-cell high rate dissolved air flotation system (supracell).

air is used. In the Oxyozosynthesis system, the dissolved gases include oxygen, ozone, carbon dioxide, and air.

The retention time in the flotation chamber is usually short about 3–5 min depending on the characteristics of process water and the performance of the flotation unit. DGF units with such short retention times can treat water, wastewater, or sludge at an overflow rate of 3.5 gpm/ft^2 for a single unit, and up to 10.5 gpm/ft^2 for triple stacked units. A comparison between a DGF clarifier and a sedimentation tank shows that (13):

a. DGF floor space requirement is only 15% of the sedimentation tank.
b. DGF volume requirement is only 5% of the sedimentation clarifier.
c. The degrees of clarification of a DGF is similar to that of a sedimentation tank using the same flocculating chemicals.
d. The operational cost of the DGF clarifier is slightly higher than that for the sedimentation unit, which is offset by the considerably lower cost for financing the installation.
e. DGF clarifiers are usually prefabricated using stainless steel. This results with lower erection cost, better flexibility in construction, and ease of possible future upgrade compared with the *in situ* constructed heavy concrete sedimentation tanks.

Currently used DGF units are more reliable, have excellent performance for sludge-thickening and require less land area than gravity thickeners. However, the gas released



1. Raw water collection tank
2. Supracell feed pump
3. Supracell inlet pipe (inbedded in ground)
4. Inlet compartment
5. Settled sludge sump
6. Settled sludge discharge
7. Clarified water outlet
8. Clarified water return for level control (1)
9. Level control in supracell with pneumatic sensor and clarified water discharge regulating valve
10. Pressure pump for recycling of clarified water to the air dissolving tubes
11. Air dissolving tubes
12. Supracell main tank
13. Spiral scoop for collection of the floated sludge
14. Floated sludge
15. Steel legs
16. Second elevated supracell in steel construction

Fig. 6. A double-cell high rate dissolved air flotation system (supracell).

to the atmosphere might strip volatile organic material from the sludge. The volume of sludge requiring ultimate disposal or reuse may be reduced, although its composition will be altered if chemical flotation aids are used. US EPA data from various air flotation units indicate that solids recovery ranges from 83 to 99% at solids loading rates of 7–48 lb/ft^2/d. A summary of US EPA data that illustrate the excellent performance of DAF for thickening various types of sludges is shown in Table 3.

DAF is also an excellent process for solids separation in water treatment and wastewater reclamation (14–17). DAF is an integral part of the Oxyozosynthesis wastewater reclamation system. A bird's view of the advanced DAF unit with built-in chemical flocculation and filtration (sandfloat) is shown in Fig.7. The influent raw water or wastewater enters the inlet at the center near bottom and flows through a hydraulic rotary joint and an inlet distributor into the rapid mixing section of the slowly moving carriage. The entire moving carriage consists of rapid mixer, flocculator, air dissolving tube, backwash pump, sludge discharge scoop, and sludge recycle scoop. From the rapid mixing section, the water enters the hydraulic flocculator for gradually building up the flocs by gentle mixing. The flocculated water moves from the flocculator into the flotation tank clockwise with the same velocity as the entire carriage including the flocculator which

Table 3
Sludge-Thickening by Dissolved Air Flotation

	Feed solids conc. (%)	Loading rate without polymer ($lb/ft^2/d$)	Loading rate with polymer ($lb/ft^2/d$)	Float solids conc. (%)
Primary + WAS	2	20	60	5.5
Primary + (WAS + $FeCl_3$)	1.5	15	45	3.5
(Primary + $FeCl_3$) + WAS	1.8	15	45	4
WAS	1	10	30	3
WAS + $FeCl_3$	1	10	30	2.5
Digested primary + WAS	4	20	60	10
Digested primary + (WAS + $FeCl_3$)	4	15	45	8
Tertiary (Alum)	1	8	24	2

Source: US EPA.

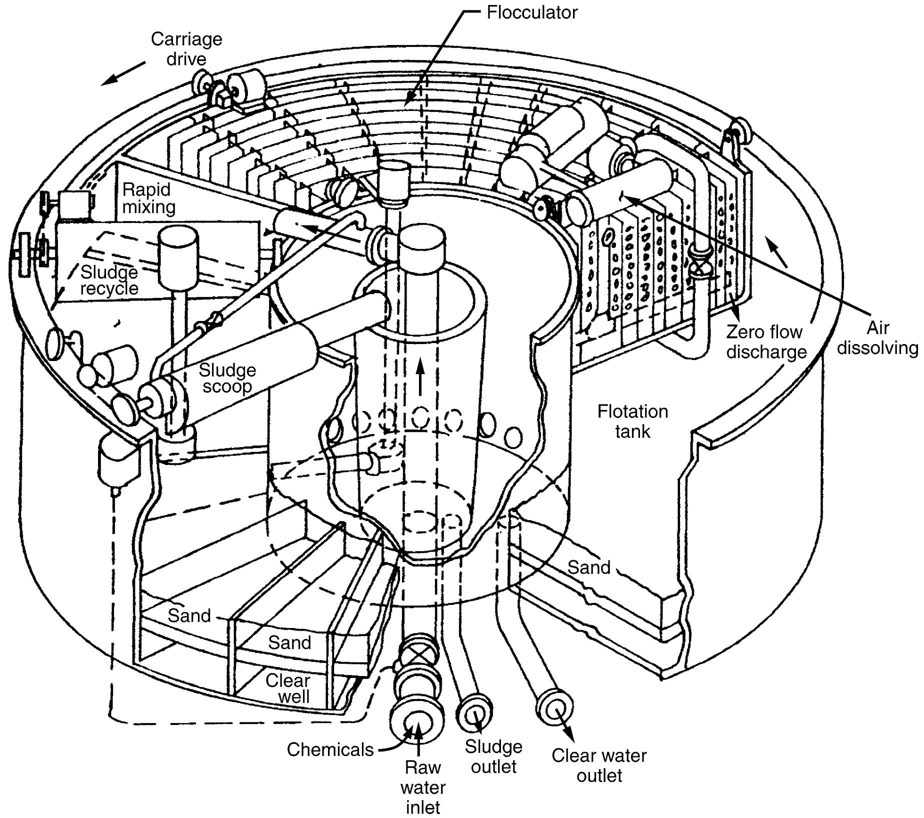


Fig. 7. Bird's view of a flocculation/flotation/filtration package unit (sandfloat).

is moving counterclockwise simultaneously. The flocculator effluent velocity is compensated by the opposite velocity of the moving carriage, resulting in a "zero" horizontal velocity of the flotation tank influent. The flocculated water thus stands still in the flotation tank for optimum clarification. At the outlet of the flocculator, clarified or recycled water stream with microscopic air bubbles is added to the flotation tank in order to float the insoluble flocs and suspended matters to the water surface. The float (scum/sludge) accumulated at the top of the unit is scooped off by a sludge discharge scoop and discharged into the center sludge collector in which there is a sludge outlet to an appropriate sludge treatment facility. The bottom of the Sandfloat is made up of multiple sections or wedges of sand filter and clear well. The clarified flotation effluent passes through the sand filter downward and enters the clear well. Through the circular hole underneath each sand filter section, the filter effluent enters the center portion of the clear well where there is an outlet for the sandfloat effluent. The filter sections are backwashed sequentially.

For the wastewater reclamation plant, DAF is an important process unit. Filtration is used for final polishing of the plant effluent. Table 4 presents the US EPA data on removal of various classical pollutants, toxic heavy metals, and toxic organics by flotation. For more information on the DAF process the reader is referred to refs. 18, 19.

2.3. Filter Belt Press

The filter belt press or simply the belt press is used for sludge-dewatering. Filter belt presses consist of an endless filter belt that runs over a drive and guide roller at each end like a conveyor belt. Several rollers support the filter belt along its length. Above the filter belt is a press belt that runs in the same direction and at the same speed; its drive roller is coupled with the drive roller of the filter belt. The press belt can be pressed on the filter belt by means of a pressure roller system whose rollers can be individually adjusted horizontally and vertically. The sludge to be dewatered is fed on the upper face of the filter belt and is continuously dewatered between the filter and press belts. After having passed the static pressure zone, further dewatering is achieved by the superimposition of shear forces to expedite the dewatering process. The supporting rollers of the filter belt and the pressure rollers of the pressure belt are adjusted in such a way that the belts and the sludge between them describe an S-shaped curve. Thus, there is a parallel displacement of the belts relative to each other as a result of the differences in the radii. After further dewatering in the shear zone, the sludge is removed by a scraper.

Some units consist of two stages in which the initial draining zone is on the top level followed by an additional lower section wherein pressing and shearing occur. A significant feature of the belt filter press is that it employs a coarse mesh, relatively open weave, metal medium fabric. This is feasible because of the rapid and complete cake formation obtainable when proper flocculation is achieved. Belt filters do not need vacuum systems and do not have the sludge pickup problem occasionally experienced with rotary vacuum filters. The belt press can handle the hard-to-dewater sludges more readily. The produced low cake moisture permits incineration of primary/secondary sludge combinations without auxiliary fuel. A large filtration area can be installed in a minimum of floor area. To avoid penetration of the filter belt by sludge, it is usually necessary to coagulate the sludge (generally with synthetic, high polymeric flocculants). However, the sludge treated by ozonation does not need any flocculants for sludge conditioning.

The process reliability is considered to be excellent. More than 1 yr of trouble-free operation has been achieved at the WNYSTP. Table 5 shows performance data collected at the WNYSTP. The last two entries in the Table represent the primary sludge at the WNYSTP and the secondary sludge that was collected from a nearby secondary treatment

Table 4
Removal of Various Pollutants, Toxic Heavy Metals, and Organics by Flotation

Pollutant	Full scale (data points)	Effluent concentration		Removal efficiency (%)	
		Range	Median	Range	Median
Classical pollutants (mg/L)					
BOD$_5$	9	140–1000	250	4–87	68
COD	12	18–3200	1200	8–96	66
TSS	12	18–740	82	6–98	88
Total phosphorus	6	<0.05–12	0.66	50 to >99	98
Total phenols	10	<0.001–23	0.66	3 to >94	12
Oil and grease	11	16–220	84	57–97	79
Toxic pollutants (μg/L)					
Antimony	9	ND to 2300	20	4–95[a]	76
Arsenic	7	ND to 18	<10	8 to >99	45
Xylene	3	ND to 1000	200	95 to >99	97
Cadmium	9	BDL to <72	3	0 to >99	98[a]
Chromium	12	2–620	200	20–99	52
Copper	12	5–960	180	9–98	75
Cyanide	7	<10–2300	54	0 to <62	10
Lead	13	ND to 1000	70	9 to >99	98
Mercury	8	BDL to 2	BDL	33–88	75
Nickel	12	ND to 270	41	29 to >99	73
Selenium	3	BDL to 8.5	2		NM
Silver	5	BDL to 66	19		45
Thallium	3	BDL to 50	14		NM
Zinc	11	ND to 53,000	200	12 to >99	89
Bis (2-ethylhexyl) phthalate	8	30–1100	100	10–98	72
Butyl benzyl phthalate	5	ND to 42	ND	97 to >99	>99
Carbon tetrachloride	3	BDL to 210	36		75
Chloroform	6	ND to 24	9	20 to >99	58
Dichlorobromomethane	1		ND		>99
2,4-Dichlorophenol	1		6		NM
Di-*N*-butyl phthalate	6	ND to 300	20	0 to >99	97
Diethyl phthalate	1		ND		>99
Di-*N*-octyl phthalate	6	ND to 33	11	61 to >99	78
N-nitrosodiphenylamine	1		620		66
N-nitroso-di-*N*-propylamine	1		84		NM
2-Chlorophenol	1		2		NM
2,4-Dimethylphenol	2	ND to 28	14		>99
Pentachlorophenol	5	5–30	13		19
Phenol	8	9–2400	71	0–80	57
2,4,6-Trichlorophenol	1		3		NM
Benzene	3	5–200	120		NM
Chlorobenzene	1		57		NM

(Continued)

Table 4 ***(Continued)***

Pollutant	Full scale (data points)	Effluent concentration		Removal efficiency (%)	
		Range	Median	Range	Median
Dichlorobenzene	2	18–260	140		76
Ethylbenzene	7	ND to 970	44	3 to >99	65
Toluene	6	ND to 2100	580	10 to >99	39
Fluoranthene	2	0.5 to <10	5.2		NM
Fluorene	1		14		NM
Naphthalene	9	ND to 840	96	33 to >99	77
Pyrene	2	0.3–18	9.2		0
Anthracene/phenanthrene	5	0.2–600	10	45 to >98	81
2-Chloronaphthelene	1		17		0

Source: US EPA.
Blanks indicate data not available.
BDL, less than detection limit.
ND, not detected.
NM, not meaningful.
[a] Approximate value.

Table 5
Belt Press Performance

Feed solids (%)	Secondary: primary ratio	Polymer dosage[a]	Pressure lb/in.^{2}g[b]	Cake solids (%)	Solids recovery (%)	Capacity[c]
9.5	100% Primary	1.6	100	41	97–99	2706
8.5	1:5	2.4	100	38	97–99	2706
7.5	1:2	2.7	25–100	33–38	95–97	1485
6.8	1:1	2.9	25	31	95	898
6.5	2:1	3.1	25	31	95	858
6.1	3:1	4.1	25	28	90–95	605
5.5	100% Secondary	5.5	25	25	95	546
5.6	100% Primary	None	NA	39–43	>97	NA
3.8	100% Secondary	None	NA	25	>95	NA

Source: US EPA.
[a] Pounds per ton dry solids.
[b] Pounds per in.^{2} (gauge).
[c] Pound dry solids per hour per meter.

plant, which were oxidized by oxygenation–ozonation before entering the belt press for dewatering.

2.4. Performance of Oxyozosynthesis Sludge Management System

The sludge management system consists of a pH adjustment unit, an innovative reactor for oxygenation–ozonation under moderate pressure (40–60 psi), DGF for sludge-thickening, and an advanced filter belt press for sludge-dewatering. The system's

Table 6
Heavy Metal Contents of Dewatered Filter-Belt-Press Cake[a]

Heavy metals mg/kg dry sludge	West NY sludge cake	NJ DEP limits for land application	US EPA ceiling limits[b] for land application	US EPA high-quality limits[c] for land application
Cadmium	3	25	85	39
Chromium	14	1000	3000	1200
Copper	447	1000	4300	1500
Nickel	9	200	420	420
Lead	126	1000	840	300
Zinc	192	2500	7500	2800

Source: US EPA.

[a] The Oxyozosynthesis system hyperbaric unit was operated at pH 4.0 and contact time 90 min.

[b] Absolute value of any single concentration (40 CFR part 503 regulations, US EPA, 1994).

[c] Monthly average values (40 CFR part 503 regulations, US EPA) (23).

Table 7
Toxic Organic Compounds in Dewatered Filter-Belt-Press Cake[a,b]

Toxic organics (mg/kg dry sludge)[c]	US EPA limitations	WNY dewatered sludge cake
Aldrin	0.10	<0.001
Chlordane	0.10	<0.001
Dieldrin	0.10	<0.001
Endrin	0.10	<0.001
Heptachlor	0.10	<0.001
Heptachlor epoxide	0.10	<0.001
Lindane	0.10	<0.001
Methoxychlor	0.25	<0.001
Mirex	0.25	<0.001
p, p′—DDT	0.25	<0.001
p, p′—DDE	0.25	<0.001
p, p′— TDE (DDD)	0.25	<0.001
Toxaphene	1	<0.001
PCB	0.50	<0.001

Source: US EPA.

[a] West NewYork sewage treatment plant and US EPA.

[b] Oxyozosynthesis process' hyperbaric unit was operated at pH, 4.0; detention time, 90 min.

[c] 1 mg/kg dry sludge = 1 ppm on dry weight basis.

overall mechanical reliability is excellent. Tables 6 and 7 document the operational data at the WNYSTP (20). It is shown that the resulting cake is low in heavy metals and toxic organics, and meets the requirements of the US EPA (40 CFR part 503 regulations) (21) and the NJ Department of Environmental Protection for sludge disposal. The ozone-treated sludge cake has low volatile solids content; high suspended solids consistency; high fuel value (more than 7500 Btu/lb dry sludge); and is nonoffensive, odor-free, and almost coliform-free. Besides, the ozone-treated sludge can be thickened easily by flotation and subsequently dewatered by the filter belt press without any additional chemicals. The product sludge cake can be disposed of safely in a sanitary landfill site, spread on land for crop production, or could be reused as an ideal refuse-derived fuel (RDF).

The flotation unit uses the pressurized gases in the hyperbaric reactor vessel for water sludge separation. The pressurized gases include oxygen, ozone, and carbon dioxide. Under optimum operation, all gaseous ozone should disappear and the flotation process should release mainly oxygen and carbon dioxide. Because supplemental air is not needed in sludge flotation, a significant cost-saving in sludge-thickening is achieved.

The side streams from the flotation unit and belt press, which contain low concentrations of suspended solids and no harmful microorganisms, are recycled to the head of the treatment plant for reprocessing. The suspended solids, BOD, COD, and total Kjeldahl nitrogen (TKN) of the recycle liquors are significantly lower than that produced from aerobic digestion, anaerobic digestion, and thermal treatment processes. Therefore, there will not be any adverse effect on the biological wastewater treatment system if the side streams are recycled. pH adjustment may be needed if the ratio of low pH recycle liquor flow to the plant influent flow is high.

The heavy metal content in the recycle liquors will not be high if the wastewater treatment plant treats only the municipal sewage. In industrial areas where heavy metals could settle with the sludge by chemical precipitation or biological assimilation, many of these heavy metals will become soluble and present in the recycle liquor if the pH of influent sludge is to be lowered to 3–4 before entering the hyperbaric reactor for oxidation. In this case, two remedies are possible:

a. Operating the hyperbaric reactor without acidification. This implies a lower ozonation efficiency.
b. Operating the flotation unit with chemical additions for both pH adjustment and heavy metals flotation. This is the perfect solution for removing the heavy metals and maintaining high ozonation efficiency in the hyperbaric reactor.

In summation, the Oxyozosynthesis sludge management system is a very promising and sound engineering development (22). It will be extremely competitive under the following conditions:

a. Ocean dumping is not allowed, which is the case in the US.
b. Federal and state regulations for disposal of sludge on land are very stringent whereby the treated sludge must be stabilized and rendered safe for cropland disposal.
c. Incineration, which creates some air pollution, is not allowed in urban areas with many high-rise buildings.
d. Wet air oxidation, which creates some odor problems, is not allowed in urban areas or cannot be afforded in rural areas.
e. Distance is too far to transport sludge to another plant or site for disposal.
f. There are engineering demonstration grants available encouraging testing and/or using innovative sludge management technology.

2.5. Performance of Oxyozosynthesis Wastewater Reclamation System

The major components of the Oxyozosynthesis wastewater reclamation system (*see* Fig. 4) are two hyperbaric oxygenation–ozonation reactors (*see* Fig. 3) and a sandfloat

Table 8
Water Quality Criteria for Reclaimed Water Use in Apartment Complexes

Item	Unit	Criteria
Odor	–	Nonexistence
Color	Unit	<10
Turbidity	Unit	<5
TDS	mg/L	<1000
SS	mg/L	<5
pH	Unit	5.8–8.6
COD	mg/L	<20
BOD_5	mg/L	<10
PO_4^{3-}	mg/L	<1
MBAS	mg/L	<1
Coliform	Count/mL	Nonexistence
General bacteria	Count/mL	<100
Residual chlorine	mg/L	>0.2
TOC	mg/L	<15

flotation–filtration package unit (*see* Fig. 7). The full-scale hyperbaric reactors have a capacity of 22,000 gpd (20,23). The package unit consists of chemical flocculation, DGF, and rapid sand filtration with a full-scale plant capacity of 1 MGD that was installed in the Town of Lenox, MA for potable water treatment (24).

The aim of this combined system is to convert municipal wastewater to reusable water, meeting the water quality criteria indicated in Table 8 for reclaimed water reuse in apartment complexes (25). The ultimate goal is to renovate wastewater for reuse as a potable water supply that meets the US EPA Drinking Water Standards (26).

3. FORMATION AND GENERATION OF OZONE

3.1. Formation of Ozone

The conversion of oxygen (O_2) into ozone (O_3) requires the rupture of the very stable O_2 molecules. Because the breaking of the oxygen–oxygen bond requires a great deal of energy, only very energetic processes can accomplish it. In an electric discharge through an oxygen stream, collisions occur between electrons and oxygen molecules. A certain fraction of these collisions occur when the electrons have sufficient kinetic energy to dissociate the oxygen molecule,

$$O_2 + e^- \rightarrow 2(O) + e^- \tag{1}$$

Each of the oxygen atoms may subsequently form a molecule of ozone,

$$(O) + 2O_2 \rightarrow O_3 + O_2 \tag{2}$$

Collisions capable of dissociating oxygen molecules also occur when oxygen is bombarded with a high-speed α- or β-particles coming from radioactive processes or with the cathode rays brought out through the thin metal foil window of the Coolidge tube. The dissociation of oxygen, with subsequent formation of ozone, may also be brought

Table 9
Energy Yield and Maximum Ozone Concentration Attainable by Various Generation Methods

Methods	Energy yield (g/kWh)	Ozone concentration
Electric discharge in oxygen	Up to 150	Up to 6 vol %
Electrolysis of water	Up to 12	Up to 20%
Photochemical		
1850–2537 Å	Up to 25	Up to 0.25%
1400–1700 Å		Up to 3.5%
Radiochemical		
Using O_2 gas	220	60 ppm
Using liquid O_2	108	5 mole %
Thermal	56	0.33 mole %

about by the absorption of ultraviolet (UV) (150–190 nm) or γ-radiation, or even thermal dissociation; for instance if oxygen which has just been heated to a very high temperature (>3000°C) is suddenly quenched with liquid oxygen, a certain amount of ozone is found.

The energetic processes necessary for producing ozone molecules are also capable of destroying them. Ozone can be dissociated according to Eq. (3):

$$O_3 \rightarrow O_2 + (O) \tag{3}$$

This would not matter, of course, if it was always formed again by reaction of Eq. (2). Unfortunately there is another reaction,

$$(O) + O_3 \rightarrow 2O_2 \tag{4}$$

The higher the ozone concentration, the higher the rate for ozone destruction, so that whatever method is used for producing ozone, the concentration cannot be increased beyond a limiting value, at which the rates of formation and destruction are equal.

Ozone can also be made from water by electrolysis. Under special conditions (high current density, low temperature, adding the correct amount of sulfuric or perchloric acid to the water, and so on) the anode gases might consist of a mixture of oxygen and ozone. The reaction shown as Eq. (5) is more endothermic (207.5 kcal) than the reaction shown as Eq. (6) (34.1 kcal); therefore, it is difficult to carry out and poor ozone yields are usually obtained:

$$3H_2O \rightarrow O_3 + 3H_2 \tag{5}$$

$$3O_2 \rightarrow 2O_3 \tag{6}$$

The yields and maximum concentrations attainable by these different processes vary considerably, as seen in Table 9. It should be noted that maximum energy yields could only be obtained by operating ozone generation at lesser than the maximum ozone concentrations.

3.2. Generation of Ozone

The two technologies for generating ozone that have found practical application are the silent electric discharge and the photochemical methods. The latter is only used

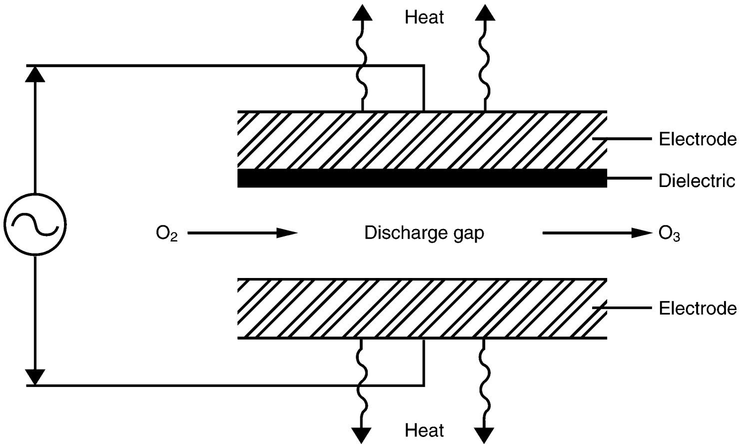


Fig. 8. Cross-section view of principal elements of a corona discharge ozone generator (*Source:* US EPA).

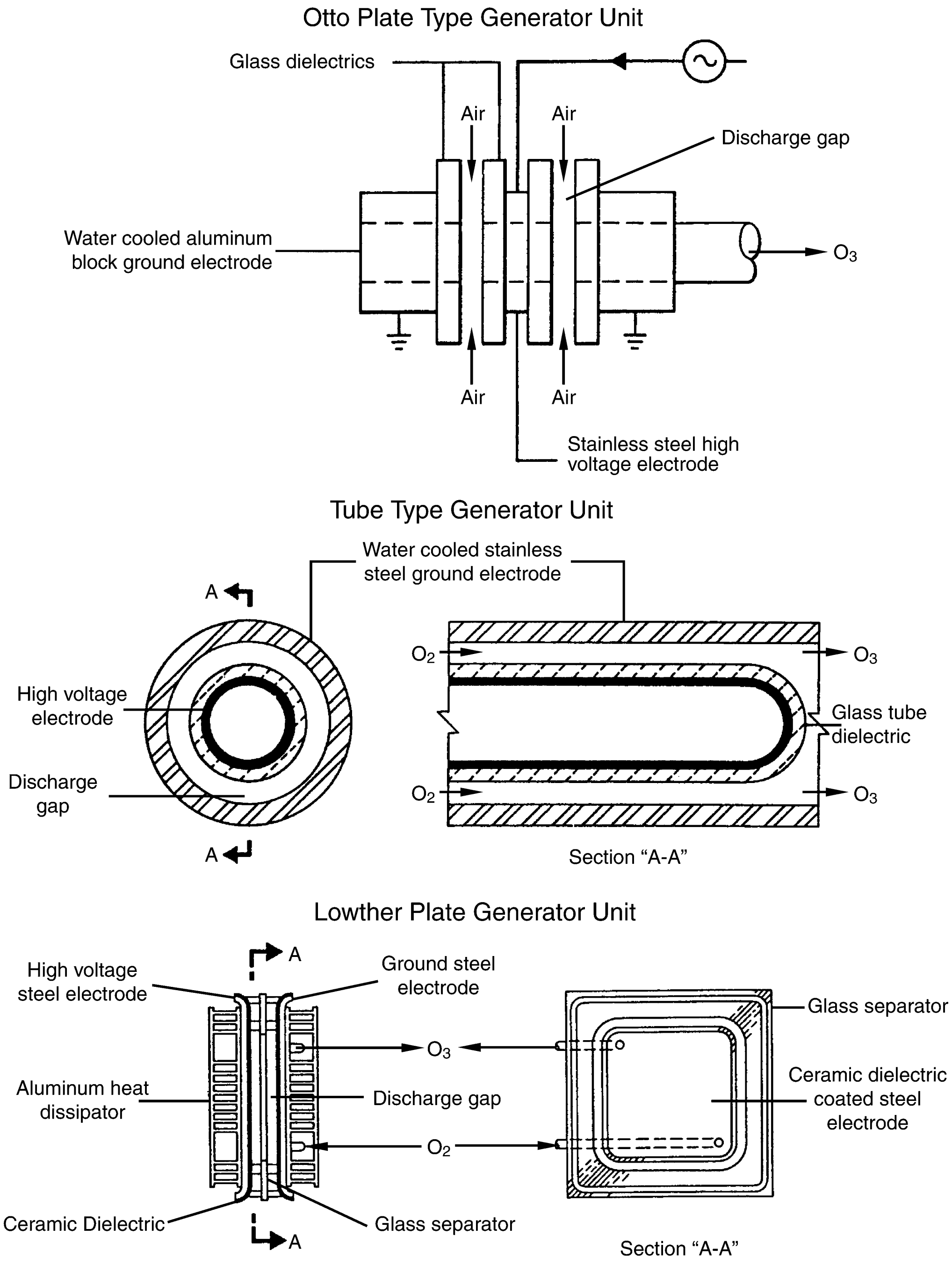
where small quantities of ozone and very low concentrations are desired. Practically, in all other laboratories and in industrial applications, the electric discharge method is used. The instability of ozone with respect to decomposition back to oxygen dictates the need for an on-site production facility. This in turn dictates the need for a cost efficient, space efficient, low maintenance installation if ozone is to be applied in wastewater and/or sludge-treatment applications. In recent years, great strides have been taken in providing equipment and technology for such installations (27–30).

Figure 8 shows the principal elements of a corona discharge ozone generator (31,32). A pair of large-area electrodes is separated by a dielectric about 1–3 mm in thickness and an air discharge gap approx 3 mm wide. When an alternating current (AC) is applied across the discharge gap with voltages between 5 and 25 KV in the presence of an oxygen containing gas, a portion of the oxygen is converted to ozone.

The excitation and acceleration of stray electrons within the high-voltage AC field cause the electrons to be attracted first to one electrode and then to the other. At sufficient velocity, these electrons split some oxygen molecules into free radical oxygen atoms, as shown in Eq. (1). The free radical oxygen atoms then combine with other oxygen molecules to form ozone according to Eq. (2).

The decomposition of ozone back to oxygen, Eq. (3) is accelerated with increasing temperature and moisture so that all generators must have a cooling device for heat removal and a drying device for moisture removal from the feed gas. For optimization of ozone generation, the following practical engineering requirements should be met:

a. For prevention of ozone decomposition, heat removal should be as efficient as possible.
b. For dielectric material and electrode protection the gap should be constructed so that the voltage can be kept relatively low while maintaining reasonable operating pressures.
c. For high-yield efficiency, a thin dielectric material with a high dielectric constant, such as glass, should be used.
d. For prolonged generator life and reduced maintenance problems, high frequency AC should be used. High frequency is less damaging to the dielectric surfaces than high voltage.



Otto Plate Type Generator Unit

Tube Type Generator Unit

Lowther Plate Generator Unit

Fig. 9. Types of ozone generators.

There are three basic types of commercial ozone generators (refer to Fig. 9). The characteristics and power requirements for the generators are given in Table 10. In addition to the generator's ozone yield per unit area of electrode surface, the concentration of ozone from the generator is regulated by:

Table 10
Comparison of Conventional Ozone Generators (Ozonators)

Typical ozonator operating characteristics	Type of ozonator		
	Otto	Tube	Lowther
Feed	Air	Air, oxygen	Air, oxygen
Dew point of feed (°F)	–60	–60	–40
Cooling	Water	Water	Air
Pressure	0	3–15	1–12
Discharge gap (in.)	0.125	0.10	0.05
Voltage (kV peak)	7.5–20	15–19	8–10
Frequency (Hz)	50–500	60	2000
Dielectric thickness (in.)	0.12–0.19	0.10	0.02
Power requirements[a]			
Air feed	10.2	7.5–10	6.3–8.8
Oxygen feed	–	3.75–5	2.5–3.5

[a] *Note*: kWh/lb of ozone at 1% concentration.

a. Adjusting the flow rate of feed gas.
b. Adjusting the voltage across the electrodes and/or
c. Selecting a suitable feed gas.

For reasons of economics, it is advisable to feed oxygen or oxygen enriched air (instead of ordinary air) to the ozone generators. However, for an electronic ozone generator using the latest semiconductors for power generation and titanium oxide ceramic electrodes for ozone generation, feeding ordinary air is common. This type of generators can deliver an ozone concentration of 2% by weight from predried air at 4.5 kWh/pound of ozone. This new ozone generation technology renders the cost of ozonation competitive with the cost of chlorine oxidation. Table 11 presents some comparative data in ozone technology. It is important to recognize that the low operating voltage (6.5 kV) of the titanium oxide ceramic electrode insures longer life and minimum maintenance.

4. REQUIREMENTS FOR OZONATION EQUIPMENT

Basically, an ozonation system consists of (33):

a. Feed gas equipment.
b. Ozone generators.
c. Ozone contactors.

4.1. Feed Gas Equipment

Conventional ozone generators are fed either with predried air or pure oxygen. The reason for the use of pure oxygen is primarily to increase the ozone concentration from 1 to 2% by weight. This factor represents a two to three times higher sterilizing and oxidative power. Because new electronic generators do not have any appreciable gain when fed with pure oxygen, it is therefore recommended that only predried air be used.

Table 11
Comparative Data in Ozone Technology

Electronic ozone generators	Conventional ozone generators
Air preparation	
Oilless compressor and heatless air dryer	Refrigerated
Dryness of air: –60°F dew point	Dryness of air: –60 to –40°F dew point
Ozone production in relation to dryness of air: 98%	Ozone production in relation to dryness of air: 70-85%
Air requirements per lb of ozone	
10.69 scfm ct 80 psi	20 scfm
Power requirements per lb of ozone	
4.035 kWh	10–12 kWh
Energy saving per lb of ozone	
Air: 2.13 kWh	None
Power: 6.965 kWh	None
Total: 9.095 kWh	
Ozone concentration from predried air	
1.6–2% by weight which represents two to three times higher sterilizing and oxidative power as compared with 1%	Maximum of 1% on the average
Ozone producing electrodes	
Material: titanium oxide ceramic	Material: glass
Dielectric strength: $e = 85$	Dielectric strength: e = 25
Dielectric constant: >15 kV/mm	Dielectric constant: <10 kV/mm
Operating voltage	
6500 V	12–16,000 V on the average
Probable failure in relationship to high voltage	
0.35%	8%
Physical size of ozone generator	
19 lb/d/30 ft^3. Weight: 330 lbs	19 lb/d/60 ft^3. Weight: 2000 lbs

Source: US Ozonair Corp.

For air preparation, equipments are required for air compression, air filtration, and air-drying:

a. **Air compression.** Oil-free compressor should be specified. More than 15 screw-type compressors are recommended as a result of their extended life. The compressor rating should be up to 100 psi.
b. **Air filtration.** Prefilter, after-filter, and after-cooler are integral parts to be supplied and mounted on the compressor. Smaller size compressors up to 5 hp are mounted on air receiver tank of appropriate size.
c. **Air drying.** Predried air at –60°F dew point is required in order to deliver 98% of the rated ozone capacity. Refrigerated or heated air dryers are capable of delivering a maximum of –50°F dew point and they are subject to failure. Only heatless air-dryers should be specified. A pressure regulator is required to control an appropriate pressure for the ozone generator in the range of 10–20 psi. A moisture indicator (colometric) should be mounted after the air dryer. The required amount of air is usually based on a maximum flow rate of 10.7 ft^3/min/lb of produced ozone.

4.2. Ozone Generators

Previously, the specifications called for conventional ozone generators to have glass electrodes with transformers rated at 16,000 V. Many design engineers have specified two identical ozone generators (one as standby) especially for larger installations. New ozone generators are being designed for a constant ozone production and constant ozone concentration. Independently wired modules control the ozone output. The specifications are currently written along the following lines (34–36):

a. **Ozone generator.** The capacity is specified as weight of ozone per unit time such as lb/h or lb/d (kg/h or kg/d).
b. **Ozone concentration from predried air.** Minimum 1.6% by weight.
c. **Air requirements at −60°F dew point.** Maximum 10.7 ft^3/min/lb of ozone. Air pressure supplied to generator is 15 psi.
d. **Overall design.** Modular, wherein each module is wired and controlled from the front panel or by remote control.
e. **Power consumption.** Not more than 4.5 kWh/pound of ozone produced.
f. **Power requirements.** 220 V AC, 50 or 60 cycles.
g. **Operating voltage.** Maximum 7 kV.
h. **Control.** Front panel pushbuttons (Start/Stop); Power Indicating Light; air flow meter; DC ammeter; and AC voltmeter.
i. **Ozone resistant materials.** All parts, components, tubing, and piping in direct contact with ozone shall be of ozone resistant materials.

4.3. Ozone Contactors

It is essential to have efficient mass-transfer of ozone into the liquid. The widely used diffuser system can transfer a maximum of 65–70% of the ozone into solution. The balance of ozone (30–35%) is collected as an exhaust gas and burned. Several recently developed ozone contactors are being marketed with ozone transfer efficiency of more than 95% (37,38).

a. **In-line contactor for water treatment (Fig. 10).** The in-line contactor consists of venture-type ejector and two or more built-in static mixers. It is mounted directly in the water supply line. One or several contactors might be used and grouped into a single manifold. The maximum diameter of the contactor is 3 in. with a minimum of 40 ft of pipe run after the contactor. A minimum water pressure of 30 psi is required upstream of the contactor in order to accommodate the 40% pressure loss in the contactor itself. Depending on the degree of contamination, it is possible in some instances to supersaturate with dissolved ozone only part of the flow and then mix it with the untreated water flow.
b. **Film layer purifying chamber (FLPC) contactor for water treatment (Fig. 11).** The basic principle is the reverse of that for bubbling. Contaminated water is emulsified (sprayed) into a powerful ozone concentration. The net result is that up to 1.5 mg/L of ozone is dissolved almost instantly in the water (compared with 0.5 mg/L concentration in ozone bubbling). FLPC-treated water is discharged under gravity following a retention period of 2–4 min in the tank under an influent water pressure of 30 psi.
c. **Turbine contactor for wastewater treatment (Fig. 12).** The turbine contactor is used for wastewater treatment in which the ozone contact time has to be extended. Mass transfer of 10–12 lb of oxygen per horsepower as compared with 2–3 lb with an average aerator. Ozone transfer of 99–100% can be achieved. For efficient wastewater treatment, one to four turbines might be used in an oxidation ditch depending on the effluent contamination and

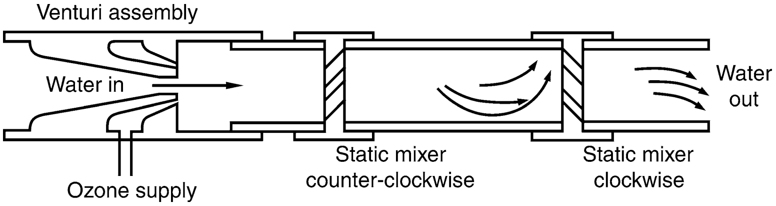


Fig. 10. In-line ozone contactor.

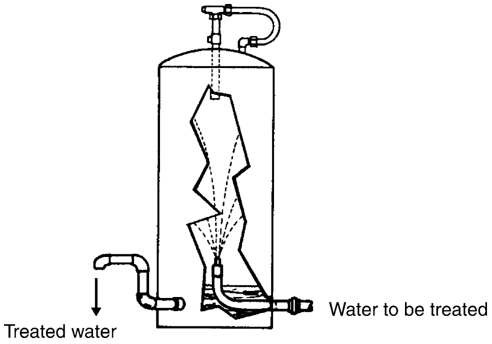


Fig. 11. Film layer purifying chamber (FLPC) ozone contactor.

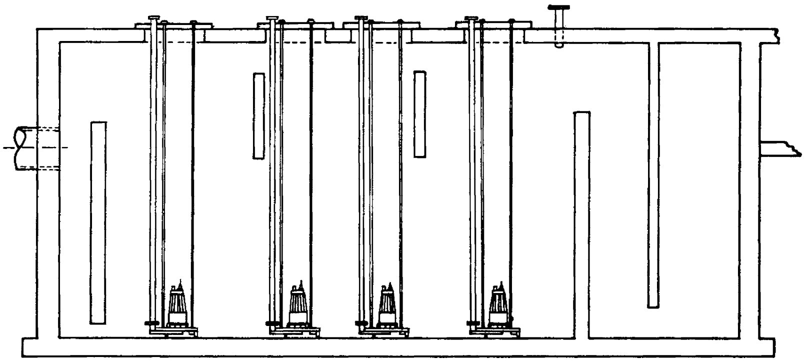


Fig. 12. Multicompartment turbine ozone contactor.

the flow. A multicompartment contactor where the effluent is introduced into fresh ozone residual, approx 20% higher treatment effect will be obtained compared with treating the effluent in one single compartment.

d. **Diffuser contactor for water and wastewater treatment (Fig. 13).** Disinfection and some chemical oxidation processes are mass-transfer-rate limited, whereas others are chemical-reaction-rate limited. Diffuser contactors are designed as part of an overall system to optimize the tradeoffs between ozone transfer and the contact time required for achieving a specific treatment objective. Systems are designed for a minimum ozone transfer of 90% and a typical disinfection contact time of 15 min. Both the influent upward velocity and the effluent downward velocity should be ≤ 0.5 ft/s. Extensive pilot plant and modeling studies, considering such factors as diffuser type, size and porosity, and arrangement in relation to mass transfer, mixing, baffling, wall effects, and materials of construction, have led to the currently recommended designs.

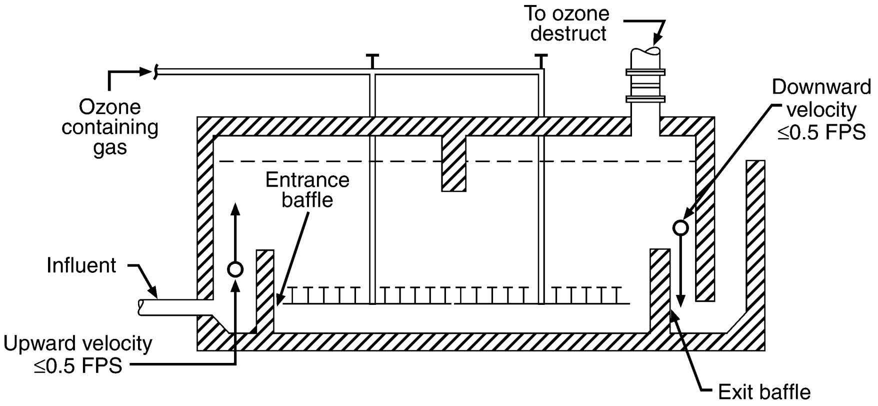


Fig. 13. Diffuser ozone contactor.

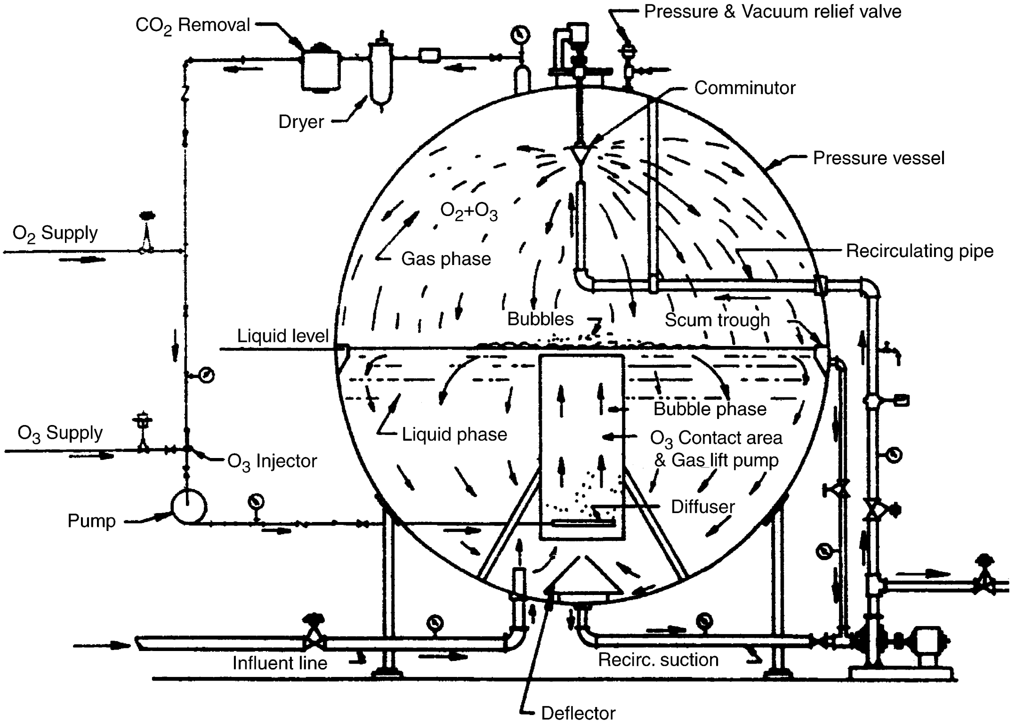


Fig. 14. Pressurized oxygen-ozone contactor (hyperbaric reactor vessel).

e. **Hyperbaric vessel for both wastewater and sludge treatment (Fig. 14).** This innovative ozone–oxygen contactor is a combination of conventional diffuser contactor, FLPC contactor, and turbine contactor. It is an advanced contactor, which is suitable for wastewater effluent and sludge treatment. Oxygen is first pumped into the hyperbaric vessel until a pressure of 40 psig is reached. Ozone is then pumped into a small compartment in the reactor through a gas diffuser, eventually making its way into the second main compartment of the reactor. Part of the wastewater or sludge is recycled by a recirculation pump and emulsified (sprayed) into the powerful ozone–oxygen concentration zone near the top of the reactor with the aid of a comminutor. A film layer is thus created for efficient gas transfer.

5. PROPERTIES OF OZONE

Ozone is an unstable colorless gas which condenses to a dark blue liquid. It has a characteristic odor to which it owes its name, derived from the Greek word “ozein,” to smell. The odor of ozone in the vicinity of an electrical machine is well known. It is generally encountered in dilute form in a mixture with air or oxygen. Ozone is formed photochemically in the earth’s stratosphere but, at ground levels, it exists only at great dilution. It is produced commercially from air or oxygen by a form of electric discharge. It is a potent germicide and powerful oxidant in both inorganic and organic reactions. With unsaturated organic compounds ozone adds to the carbon–carbon double bond, forming ozonides. Decomposition of these ozonides almost always results in cleavage at the double bond, a property, which has been used for structural analysis and in the commercial preparation of chemicals.

At ordinary temperatures and the concentrations it is normally produced, the color is not noticeable unless the gas is viewed through considerable depth. At –112ºC, ozone condenses to a dark blue liquid. Liquid ozone is easily exploded, as are concentrations of ozone–oxygen mixtures above 20% ozone, in either the liquid or the vapor state. Explosions may be initiated by minute amount of catalysts or organic matter, shocks, electric sparks, sudden changes in temperature or pressure, and so on.

Ozone has strong absorption bands in the infrared, the visible, and the UV. The absorption maximum at 253.7 nm is particularly strong and affords a convenient means of measuring ozone concentrations in the stratosphere, as well as in the laboratory and in industrial measurements. Other properties of ozone are given in Table 12.

Liquid ozone is reported to be miscible in all proportions with the following liquids: $CClF_3$, $CC1_2F_2$, CH_4, CO, F_2, NF_2 and OF_2; forms two-layer systems with the following liquids: CF_4, N_2, and O_2; and Ozone solutions in $CC1F_3$ (chlorotrifluromethane) (about 105 g/L) have been prepared commercially in small cylinders. It is necessary to refrigerate these cylinders (–75°C) to minimize the decomposition of ozone at higher temperatures.

The limited miscibility of ozone in oxygen is of practical importance because the dense, oxygen-rich layer, which settles to the bottom, is easily exploded. The mutual solubility of the two liquids decreases when the temperature is reduced. Thus, liquid ozone and oxygen are completely miscible above 93.2 K (at which temperature the total pressure is 1.25 atm) but at 90.2 K (the atmospheric-pressure boiling point of liquid O_2), there is a separation into two layers, containing 17.6 and 67.2 mol % ozone, respectively. At still lower temperatures the separation becomes even more pronounced.

Ozone gas is sparingly soluble in water, and more so in other liquids, especially at low temperatures. The solubility of ozone in water is given in Table 13.

$$B = \frac{\text{ozone concentration in liquid}}{\text{ozone concentration in gas reduced to STP}} \tag{7}$$

where B is the Bunsen coefficient of solubility; and H is the Henry’s constant, mole/atm (mole fraction of ozone in solution/partial pressure of ozone in gas, in atm).

The preparation of saturated ozone solutions is difficult to achieve because of the great tendency of ozone to react or to undergo decomposition. The thermal decomposition of

Table 12
Properties of Pure Ozone

Parameters	Data
Melting point (°C)	−192. 5 ± 0.4
Boiling point (°C)	−111. 9 ± 0.3
Critical temperature (°C)	−12.1
Critical pressure (atm)	54.6
Critical volume (cm^3/mole)	111
Density of solid ozone (g/cm^3), at 77.4 K	1.728
Viscosity of liquid (cP)	
At 77.6 K	4.17
At 90.2 K	1.56
Surface tension (dyn-cm)	
At 77.2 K	43.8
At 90.2 K	38.4
Parachor[a]	
At 90.2 K	75.7
Dielectric constant (liquid)	
At 90.2 K	4.79
Dipole moment, debye	0.55
Magnetic susceptibility (cgs units)	
Gas	0.002×10^{-8}
Liquid	0.150
Heat capacity of liquid from 90 to 150 K	$C_p = 0.425 + 0.0014\,(T–90)$
Heat of vaporization, kcal/mole,	
At −111.9°C	3410
At −183°C	3650

Density and vapor pressure of liquid

Temperature (°C)	Density (g/cm^3)	Vapor pressure (torr)
−183	1.574	0.11
−180	1.566	0.21
−170	1.535	1.41
−160	1.504	6.73
−150	1.473	24.8
−140	1.442	74.2
−130	1.410	190
−120	1.378	427
−110	1.347	865
−100	1.316	1605

Heat and free energy of formation	ΔH_f (kcal/mole)	ΔG_f (kcal/mole)
Gas at 298.15 K	34.15	38.89
Liquid at 90.15 K	30	–
Hypothetical gas at 0 K	34.74	–

[a] $M\gamma^{1/4}(D–d)$ where M, molecular weight; γ, surface tension; D, liquid density; d, vapor density.

ozone has been extensively studied in the temperature range 80–500°C. The mechanism is as shown in Eqs. (8)–(10).

$$O_3 + M \rightarrow O_2 + (O) + M \tag{8}$$

$$O_2 + (O) + M \leftrightarrows O_3 + M \tag{9a,b}$$

Table 13
Solubility of Ozone in Water

Temperature (°C)	Bunsen coefficient	Henry's law coefficient ($H \times 10^{-4}$)
0	0.49	3.95
5	0.44	3.55
10	0.375	3
20	0.285	2.29
30	0.20	1.61
40	0.145	1.17
50	0.105	0.85
60	0.08	0.64

Table 14
Uncatalyzed Thermal Decomposition of Ozone in Ozone–Oxygen Mixtures[a]

			Half-life for indicated initial concentration[b]			
		Wt %	5	2	1	0.5
Temp (°C)	$k_2/2k_1k_3$	Vol %	3.333	1.333	0.667	0.333
120	22.4		11.2[c]	28[c]	56[c]	112[c]
150	1.40		41.8	104.5	209	418
200	0.030		0.9	2.2	4.5	9
250	0.00133		0.04	0.10	0.20	0.40

[a] *See* ref. 39.
[b] Half-life = $(k_2/2k_1k_3)$ (100/vol % O_3) in seconds.
[c] In minutes.

$$(O) + O_3 \rightarrow 2O_2 \qquad (10)$$

where *M* is a third element, O_2, O_3, N_2, He, or whatever is present. This leads to the rate expression shown in Eq. (11).

$$-d(O_3)/dt = 2\,k_1k_3(O_3)^2/k_2O_2 \qquad (11)$$

where k_1, k_2, and k_3 are rate constants. Because the thermal decomposition of ozone is not a first order process, the half-life of the ozone varies inversely with its initial concentration and directly with the oxygen concentration. Typical values are given in Table 14 (39).

Numerous substances can catalyze the decomposition of ozone. The reaction with N_2O_5 proceeds according to the mechanism shown in Eqs. (12)–(15).

$$N_2O_5 \rightarrow NO_2 + NO_3 \qquad (12)$$

$$NO_2 + NO_3 \rightarrow N_2O_5 \qquad (13)$$

$$NO_2 + O_3 \rightarrow NO_3 + O_2 \qquad (14)$$

$$2NO_3 + 2NO_2 \rightarrow 4NO_2 + O_2 \qquad (15)$$

As long as any ozone remains, the N_2O_5 is regenerated so that the net effect is the decomposition of ozone. This process was studied at 20° and 400°C, and might be of some importance in the case of ozone generated from air because air always contains

Table 15
Comparison of the Solubilities of Ozone, Chlorine, and Oxygen

	Solubility by water temperature (mg/L)			
Gas	0°C	10°C	20°C	30°C
Oxygen				
At 100%	70.5	54.9	44.9	38.2
At 21%	14.8	11.5	9.4	8
Ozone				
At 100%	1374.3	1114.9	789	499.6
At 4%	55	44.6	31.6	20
Chlorine				
At 100%	14,816.5	9963.4	7263.6	5688.8
At 99.8%	14,789.4	9943.5	7249.1	5677.4

traces of N_2O_5. At room temperature, the decomposition of ozone apparently depends on surface reactions. A half-life of 20–100 h may be expected in clean vessels of glass, stainless steel, or other inert materials. Many solids catalyze the decomposition of ozone. The activity of such catalysts depends on subdivision, crystal structure, presence or absence of moisture, and so on. Preparations of iron oxide have been made that are extremely active in decomposing ozone.

In aqueous solutions, the decomposition of ozone is much more rapid than in the gaseous state. It is catalyzed by the hydroxyl ion. The initial reaction is shown in Eq. (16),

$$O_3 + OH^- \rightarrow O_2 + (HO_2) \tag{16}$$

followed by the reactions shown by Eqs. (17)–(20).

$$O_3 + (HO_2\ ) \rightarrow 2O_2 + OH^- \tag{17}$$

$$O_3 + OH^- \rightarrow O_2 + (HO_2) \tag{18}$$

$$2(HO_2) \rightarrow O_3 + H_2O \tag{19}$$

$$(HO_2) + (HO) \rightarrow O_2 + H_2O \tag{20}$$

Ozone is more soluble in water than is oxygen, but, because of a much lower available partial pressure, it is difficult to obtain a concentration of more than a few milligrams per liter under normal conditions of temperature and pressure.

A comparison of the solubilities of ozone, chlorine, and oxygen by water temperature and gas concentration is presented in Table 15. A mathematical model describing DO concentration can be found elsewhere (40). The DO concentration is a function of water temperature, pressure, and chloride concentration.

In 1981, Hill et al. (41) performed ozone absorption in a pressurized bubble column (7.7 m tall and 5.25 cm inside diameter), which was operated in a semibatch mode with gas pressures up to 791 kPa (100 psig) and water temperatures ranging from 20 to 40°C. Also in 1981, Roth and Sullivan (42) reviewed and investigated the solubility of ozone in water under various pH values and water temperatures. Their reviewed data

Table 16
Ozone Solubility in water[a]

Investigator	Temp (°C)	H (atm/mole fraction)
Kawamura (1932)	5	2880
	10	3400
	20	4610
	30	6910
	40	9520
	50	13,390
	60	18,980
Kawamura (1932)		
7.57 N H_2SO_4	20	7420
2.02 N H_2SO_4	20	5810
1.01 N H_2SO_4	20	5350
0.18 N H_2SO_4	20	4770
0.11 N H_2SO_4	20	4770
Kirk-Othmer (1967)	0	2530
	5	2820
	10	3330
	20	4370
	30	6210
	40	8550
	50	11,770
Li (1977)		
pH 2.2	25	7840
pH 4.1	25	7600
pH 6.15	25	9000
pH 7.1	25	9400

[a] *See* ref. 42.
See refs. 107–109.

are presented in Table 16 and Eq. (21) is a mathematical model fitting their experimental data:

$$H = 3.84 \times 10^7 \, (\mathrm{OH}^-)^{0.035} \exp(-2428/T) \qquad (21)$$

where H is the the Henry's Law constant (atm/mole fraction of ozone); (OH^-) is the hydroxide concentration (g-mole/L); and T is the temperature (°K).

Ozone supposedly decomposes in water, but this is probably because of its strong oxidizing ability rather than simple decomposition. Ozone is much more soluble in acetic acid, acetic anhydride, dichloroacetic acid, chloroform, and carbon tetrachloride than it is in water. More technical information on ozonation can be found elsewhere (43–62).

It should be noted that HO_2 and HO in Eqs. (16)–(20) are free radicals, which are formed when ozone decomposes in aqueous solutions. The two free radicals have great oxidizing power, and in addition to disappearing rapidly Eq. (20), may react with impurities or pollutions present in solution, such as metal salts, organic substance, hydrogen, hydroxide ions, and so on. It is believed that the free radicals formed by the decomposition of ozone in water are the principal reacting species (58).

6. DISINFECTION BY OZONE

Smith and Bodkin (43) compared the bactericidal action of ozone and chlorine at varying pH values. Ozone, over a wide pH range, was many times as effective as chlorine. At a temperature of 27.5°C and pH values of 5.0 and 6.0, ozone affected sterility of a 1-L sample containing 8 × 10^5 bacteria/mL in 5 min. At pH 7.0, 8.0, and 9.0, the sterilization time was 7.5 min. The ozone concentration varied from 0.13 to 0.20 mg/L. In contrast, the concentration of chlorine required to sterilize as rapidly as ozone varied from 2.7 mg/L at pH 5.0, to 7.9 mg/L at pH 8.0.

Leiguarda et al. (44) presented a detailed account of the bactericidal action of ozone in both pure and river waters. Varying amounts of ozone were added to pure water free from ozone demand, and the water was inoculated with dilute suspensions of *Eschericia coli* or *Clostridium perfringens*. Samples were taken at various time intervals to determine the concentration of ozone and the number of bacteria present. The effects of temperature and pH on bactericidal action were also investigated. The results indicated that in water initially containing 10^4 *E. coli*/mL and 0.12 mg/L ozone, at pH 6.0 and maintained at a temperature of 10°C, no viable bacteria were found after 5 min; the ozone content had decreased to 0.09 mg/L. At pH 8.0 and higher temperatures, the efficiency was not significantly affected by temperature but was slightly higher at pH 6.0 than at pH 8.0. In addition, tests were made on the effects of ozone on the naturally occurring bacterial flora of river water. The ozone demand of this water was high, and tests were made using 1–6 mg/L of ozone. There were sizable reductions in the number of bacteria even when the amount of ozone added was insufficient to satisfy the ozone demand of the water. All organisms were destroyed when, after a contact period of 5 min, 0.08 mg/L of residual ozone was present. Similar results were obtained in experiments with river water samples that were coagulated, settled, and subsequently inoculated with *E. coli*. A total kill of vegetative forms of *C. perfringens* occurred within 5 min when 0.12 mg/L ozone was initially added to water containing 1.4 × 10^4 bacteria/mL. In water containing *C. perfringens* in concentrations of 2 × 10^3 spores/mL, at a pH of 6.0 and maintained at a temperature of 24°C, no viable spores were found after a contact period of 15 min with 0.25 mg/L of ozone, or after 2 min with 5 mg/L. At pH 8.0, bactericidal efficiency was reduced. Spores were not affected by 0.25 mg/L of ozone even after 120 min.

In a symposium on the sterilization of water, Whitson (45) reported that ozonation affected microorganism removal and improved water filterability, color, taste, and odor. Bernier (46), while comparing the bactericidal efficiencies of chlorine and ozone, asserted that ozone was the far superior disinfectant, being considerably faster than chlorine and not as notably affected by external factors such as pH and temperature.

Bringman (47) observed that 0.1 mg/L of active chlorine required 4 h to kill 6 × 10^4 *E. coli* cells in water, whereas, 0.1 mg/L of ozone required only 5 s. When the temperature was raised from 22 to 37°C, the ozone inactivation time decreased from 5 to 0.5 s. Wuhrmann and Meyrath (48) carried an investigation of the kinetics of ozone disinfection. During each experiment, the ozone concentration was kept constant by continuously bubbling air-ozone through the test solution. The results indicated that ozone disinfection was mainly a function of contact time, ozone concentration, and water temperature. These investigations revealed that the contact time with ozone necessary for

99% destruction of *E. coli* was only one-seventh that observed with the same concentration of hypochlorous acid. The death rate for spores of *Bacillus* species was about 300 times higher with ozone than with chlorine.

Hann (49) presented a detailed review of the differences between chlorination and ozonation as determined by other workers. It is reported that ozone disinfection was found to be somewhat more expensive. Turbidity interfered with its use, and organic demand had to be satisfied before germicidal action was effective.

Scott and Lesher (50) postulated a mode of action of ozone on *E. coli* based on experimental data. The primary attack of ozone was thought to be on the cell wall or membrane of bacteria, probably by reaction with the double bonds of lipids, and that cell lysis depended on the extent of that reaction. Bringman (47) reported that the mode of action of ozone differed from that of chlorine. He concluded that chlorine selectively attacked and destroyed certain enzymes, whereas, ozone acted as a general protoplasmic oxidant. Christensen and Giese (51) suggest that the primary locus of activity of ozone was the bacterial cell surface. Barron (52) hypothesized that the primary bactericidal activity of ozone was the oxidation of sulfhydril groups on enzymes. Murray et al. (53) at the University of Western Ontario, recognizing that the outermost layer of gram-negative organisms is a lipoprotein followed by a lipopolysaccharide layer, surmised that these layers would be first subject to attack by ozone. They concluded that the attack by ozone on the cell wall results in a change in cell permeability eventually leading to lysis.

Smith (54) stated that under experimental conditions where there was less than 1% survivors of *E. coli* and *Streptococcus fecalis* after 60 s exposure to 0.8 mg/L ozone, the unsaturated fatty acids (mainly C_{16} and C_{18} monoenoic acids) of the cell lipids were oxidized in the same time interval. The lipids present in bacteria are largely confined to the cytoplasmic membrane. Thus, the mechanism of disinfection of ozone is still open to question.

At the Eastern Sewage Works in the London Borough of Redbridge, Boucher and his associates (55,56) conducted experiments on microstraining and ozonation of wastewater effluents. Using an ozone dose of 10–20 mg/L most organisms were killed, although a sterile effluent was never obtained. Chlorine followed by ozone produced better results. In his conclusion, Boucher commented: “chlorination as an additional treatment to ozonation has not produced any advantage except to destroy most of the few organisms that sometimes survive ozonation. This is not considered a sufficient advantage in view of its many known disadvantages for effluent treatment, namely, the production of chloro-derivatives, which might be toxic to fish and other aquatic life or which may produce persistent tastes, difficult to remove by subsequent waterworks treatment, and the possibility of rapid after growth of microorganisms in a receiving river and all its attendant problems.”

Huibers et al. (57) determined that ozone treatment of effluents from secondary wastewater treatment plants could provide product water, which is within the US requirements for chemical and bacteriological quality of potable water. Virtually all color, odor, and turbidity were removed. Oxygen consuming organic materials, measured as COD, were reduced to less than 15 mg/L. Bacteriological tests revealed that no living organisms remained.

In Los Angeles County, California, a well-oxidized secondary effluent was treated with ozone for disinfection (62). It was found that 50 mg/L of ozone was required to

Table 17
Relative Efficiency of Ozone Disinfection[a]

Organism	L	C
Escherichia coli	500	0.001
Streptococcus faecalis	300	0.0015
Polio virus	50	0.01
Endamoeba histolytica	5	0.1
Bacillus megatherium (spores)	15	0.03
Mycobacterium tuberculosam	100	0.005

[a] pH, 7.0; temperature, 10–15°C; *L*, lethality coefficient/(mg/L)/min; *C*, residual concentration of disinfectant (mg/L).

meet the California requirement of 2.2 total coliforms/100 mL. To achieve the 200 counts of fecal coliform per 100 mL, the ozone requirement was 10 mg/L. It was also found that the removal of suspended solids down to 1 mg/L greatly improved the efficiency of ozone treatment. This study further revealed that to achieve excellent disinfection, the COD should be less than 12 mg/L and the nitrite should be less than 0.15 mg/L. The minimum required ozonation contact time was reported to be 10 min.

The kinetics of disinfection has been investigated by Morris (63). His concept of the lethality coefficient for a given disinfectant is presented as follows:

$$L = \ln(100)/\mathrm{CT} = 4.61/\mathrm{CT} \tag{22}$$

where L is the lethality coefficient [$(\mathrm{mg/L})^{-1}\ \mathrm{min}^{-1}$]; C is the residual concentration of disinfectant (mg/L); and T is the time in minutes for 99% microorganism destruction (2-log destruction).

Table 17 indicates the effectiveness of ozone for disinfection of various bacteria and viruses at pH 7.0 and temperature of 10–15°C. Table 18 further illustrates that in comparison with various forms of chlorine, ozone is a much more powerful disinfectant against enteric bacteria, amoebic cysts, viruses, and spores by factors of 10–100. The disinfection efficiency of ozone does not seem to be affected significantly within the normal pH range of 6.0–8.5. Table 19 further presents the oxidation potentials of nine strong chemical disinfectants among which ozone has the highest oxidation potential. This makes ozone the strongest disinfectant as well as the strongest oxidizing agent. The exact effect of temperature on ozone disinfection is still unknown. However, it is known that the higher the water temperature, the lower the efficiency of ozone mass transfer, which might translate to lower disinfection efficiency.

In summation, the advantages of using ozone for disinfection include but are not limited to:

a. Ozone is a better virucide than chlorine.
b. Ozone removes color, odor, and taste (such as phenolic compounds).
c. Ozone oxidizes iron, manganese sulfides, and so on.
d. Ozone oxidizes organic impurities or pollutants in water.
e. Ozone increases the DO content in water or wastewater.

Table 18
Comparison of Lethality Coefficients for Ozone and Chlorine

Agent	Enteric bacteria	Amoebic cysts	Viruses	Spores
O_3	500	0.5	5	2
$HOC1$ as Cl_2	20	0.05	1 up	0.05
$OC1^-$ as Cl_2	0.2	0.0005	0.02	0.0005
NH_2Cl as $C1_2$	0.1	0.02	0.005	0.001

Table 19
Oxidation Potential of Chemical Disinfectants

Disinfectants	Chemical reactions	Oxidation potentials (EV)
Ozone	$O_3 + 2H^+ + 2e^- \rightarrow O_2 + H_2O$	2.07
Permanganate	$MnO_4^- + 4H^+ + 3e^- \rightarrow MnO_2 + 2H_2O$	1.67
Hypobromous acid	$HOBr + H^+ + e^- \rightarrow 0.5Br_2 + H_2O$	1.59
Chlorine dioxide	$C1O_2 + e^- \rightarrow C1O_2^-$	1.50
Hypochlorous acid	$HOC1 + H^+ + 2e^- \rightarrow Cl^- + H_2O$	1.49
Hypoiodous acid	$HOI + H^+ + e^- \rightarrow 0.5I_2 + H_2O$	1.45
Chlorine gas	$C1_2 + 2e^- \rightarrow 2C1^-$	1.36
Bromine	$Br_2 + 2e^- \rightarrow 2Br^-$	1.09
Iodine	$I_2 + 2e^- \rightarrow 2I^-$	0.54

7. OXIDATION BY OZONE

7.1. Ozone Reaction with Inorganics

Potassium ozonide, (KO_3) can be prepared at low temperatures by ozone treatment of dry KOH, the superoxide KO_2, or the metal itself dissolved in liquid ammonia as shown in Eqs. (23)–(25).

$$3O_3 + 3KOH \rightarrow KOH{\cdot}H_2O + 2O_2 + 2KO_3 \tag{23}$$

$$O_3 + KO_2 \rightarrow O_2 + KO_3 \tag{24}$$

$$O_3 + K \rightarrow KO_3 \tag{25}$$

This reaction will not occur in an aqueous medium because KO_3 immediately decomposes in the presence of water, probably by the reactions shown in Eq. (26a–c).

$$KO_3 + H_2O \rightarrow K^+ + HO_3 + OH^- \tag{26a}$$

$$HO_3 \rightarrow O_2 + [OH] \tag{26b}$$

$$4[OH] \rightarrow 2H_2O + O_2 \tag{26c}$$

The ozonides NaO_3 and $(CH_4)_4NO_3$ have also been prepared from ozone. However, they also would not be produced in an aqueous medium because of their decomposition in water, similar to that shown for KO_3. These ozonides are ionic crystals containing the O_3 ion. The reaction of ozone with ammonia has been studied in the dry gaseous state, in liquid ammonia, in carbon tetrachloride solution, and in aqueous solution. The reaction is extremely fast, and the end product is always ammonium nitrate. The reaction is shown in Eq. (27).

$$2NH_3 + 4O_3 \rightarrow NH_4NO_3 + 4O_2 + H_2O \tag{27}$$

By carrying the reaction out at very low temperatures, the formation of the red ammonium ozonide (NH_4O_3) has been demonstrated. This compound decomposes rapidly to ammonium nitrate, oxygen, and water as shown in Eq. (28).

$$4NH_4O_3 \rightarrow 2NH_4NO_3 + O_2 + 4H_2O \tag{28}$$

Ammonium salts do not react again with ozone. The reaction of ozone with the lower oxides of nitrogen NO, NO_2, N_2O_3, and N_2O_4, is extremely rapid leading to the formation of nitrogen pentoxide (N_2O_5), as shown in Eqs. (29)–(32).

$$2NO + O_3 \rightarrow N_2O_5 \tag{29}$$

$$6NO_2 + O_3 \rightarrow 3N_2O_5 \tag{30}$$

$$3N_2O_3 + 2O_3 \rightarrow 3N_2O_5 \tag{31}$$

$$3N_2O_4 + O_3 \rightarrow 3N_2O_5 \tag{32}$$

By the combined action of O_3, NO_2, and ClO_2 gases nitronium perchlorate (NO_2ClO_4), can be formed as shown by Eq. (33). This is a white solid with low vapor pressure and strong oxidizing properties.

$$3NO_2 + 3ClO_2 + 2O_3 \rightarrow 3NO_2ClO_4 \tag{33}$$

In aqueous solutions, nitrites are oxidized to nitrates as shown by Eq. (34).

$$O_3 + NO_2^- \rightarrow NO_3^- + O_2 \tag{34}$$

This reaction has been used for the quantitative determination of ozone. Ozone reacts rapidly with hydrogen sulfide (H_2S). In dry gas mixtures the only reaction product appears to be SO_2, but the amount of ozone consumed per mole of H_2S has not been established with certainty. The reaction in water solution has not been adequately studied but there are reports that colloidal sulfur is among the products.

In the gas phase, ozone and sulfur dioxide (SO_2) react only slowly. In aqueous solutions, sulfates are produced. The reaction is pH dependent and complicated by induced oxidation of sulfites by oxygen, which is generally present along with the ozone. Thus, in sodium bisulfite solutions, ozone catalyzes the reaction between bisulfite and oxygen, this effect becoming more and more pronounced, the more dilute the ozone. In sufficiently dilute ozone, as many as 60 atoms of oxygen are consumed per molecule of ozone supplied. Similar observations have been made in the oxidation of aqueous SO_2 solutions.

Ozone liberates iodine from iodide solutions. The amount of iodine liberated per mole of ozone depends on the pH, concentration, temperature, and perhaps other factors. One mole of ozone liberates 1 mole of iodine, or its equivalent, in alkaline solutions; 1.1–1.3 moles of iodine at pH 7.0, and 2.5–3 mol of iodine in concentrated HI solutions. Ozone also oxidizes bromides to bromine as shown in Eq. (35).

$$O_3 + 2Br^- + H_2O \rightarrow Br_2 + 2OH^- + O_2 \tag{35}$$

Chlorides are only oxidized in acid solutions and the reaction is quite slow. Ozone oxidizes the cyanide ion (CN^-) to cyanate (CNO^-) as shown in Eq. (36), in the first stage reaction, then the cyanate is oxidized further by ozone and converted to bicarbonate, nitrogen, and oxygen as shown in Eq. (37).

$$CN^- + O_3 \rightarrow CNO^- + O_2 \tag{36}$$

$$2CNO^- + 3O_3 + H_2O \rightarrow 2HCO^- + N_2 + 5O_2 \tag{37}$$

Thiocyanates (CNS^-) are oxidized to cyanide as indicated in Eq. (38).

$$CNS^- + 2O_3 + 2OH^- \rightarrow CN^- + SO_3^{2-} + 2O_2 + H_2O \tag{38}$$

If the addition of ozone is continued, then the cyanide ion is oxidized to cyanate and the sulfite ion is oxidized to sulfate as shown in Eq. (39).

$$CN^- + SO_3^{2-} + 2O_3 \rightarrow CNO^- + SO_4^{2-} + 2O_2 \tag{39}$$

Finally the cyanate ion (CNO^-) can be oxidized to harmless bicarbonate, nitrogen, and oxygen according to Eq. (37).

Argentic (Ag^{+2}) salts can be produced by treating Ag NO_3 solutions with ozone as shown by Eq. (40).

$$Ag^+ + 2O_3 \rightarrow Ag^{2+} + 3O_2 \tag{40}$$

Ferrous (Fe^{+2}) salts are oxidized to ferric (Fe^{3+}) salts. In dilute H_2SO_4 solution the reaction is as shown in Eq. (41).

$$2Fe^{2+} + O_3 + 2H^+ \rightarrow 2Fe^{3+} + O_2 + H_2O \tag{41}$$

The reaction for the oxidation of ferrocyanide, $Fe(CN)_6^{-4}$ to ferricyanide, $Fe(CN)_6^{3-}$ proceeds according to Eq. (42). The engineering significance of this reaction is for its usefulness in the treatment of wastewater produced by the photoprocessing industry (32).

$$2Na_4Fe(CN)_6 \cdot 10H_2O + O_3 \rightarrow 2Na_3Fe(CN)_6 + O_2 + 9H_2O + 2NaOH \tag{42}$$

This review of the reactions of ozone with inorganic pollutants in water and wastewater has failed to demonstrate that stable inorganic ozonides can be produced in aqueous solution and it is concluded that inorganic ozonides would not be a problem if ozone were used as the disinfectant for water or wastewater.

For practical water or wastewater treatment, ozone oxidizes inorganic impurities by straight chemical oxidation reaction. The examples are as follows.

a. Ozone reacts rapidly to oxidize ferrous ion (Fe^{2+}) into ferric ion (Fe^{3+}). Ferric ion can then be removed as insoluble ferric hydroxide at high pH, or can co-precipitate with phosphate ion for both iron and phosphorus removals.
b. Ozone rapidly oxidizes manganous (Mn^{2+}) into insoluble manganese dioxide or soluble permangante (MnO_4^-). The permanganate is also an oxidizing agent. Stoichiometrically, 0.94 mg/L of potassium permanganate will further oxidize 1 mg/L of ferrous ion; or 1.92 mg/L of potassium permanganate will further oxidize 1 mg/L of manganous ion. Eventually

permanganate ion also converts into insoluble manganese dioxide in the aforementioned oxidation–reduction reaction.

c. Ozone can break down organometallic complexes of both iron and manganese, which usually defy the conventional oxidation processes for iron and manganese removal from potable water.
d. Ozone readily oxidizes sulfides and sulfites to stable sulfates, and nitrites to stable nitrates.
e. Bromides and chlorides can be oxidized by ozone to bromine (Br_2) and chlorine (Cl_2), respectively, although these reactions are slow and dependent upon the concentration of reactants.
f. The oxidation of iodides (I^-) to iodine (I_2) is the basis for the standard method used in the determination of ozone concentration (64).
g. The ammonium ion (NH_4^+) is apparently not attacked under normal pH values in wastewater treatment. So there is no waste of ozone oxidizing capacity with ammonia nitrogen if present in a waste stream. However, ammonia is oxidized completely to nitrate by ozone if the wastewater pH is and remains alkaline. The molar ratio of ozone consumed per ammonia oxidized is about 12:1 (65).
h. In the process of treating photoprocessing water, silver is recovered electrolytically, then the spent bleach bath of iron ferrocyanide complexes are ozonated. Ferrocyanide in the spent liquor is oxidized to ferricyanide, which is its original form. Thus the bleach is regenerated for reuse by the photoprocessor [*see* Eq. (42)].
i. Ozone may replace chlorine in the treatment of industrial wastewaters containing cyanide. Ozone oxidation takes place in multistages according to Eqs. (36)–(39). Ozonation of cyanide ions and dyes in aqueous solutions is documented in refs. 66, 67.

7.2. Ozone Reaction with Organic Material

The exact mechanism for the chemical processes taking place when water or wastewater is subjected to treatment by ozone and oxygen is still under investigation. In general, part of ozone in the reactor oxidizes carbonaceous substances in water or wastewater, thus producing carbon dioxide and oxygen, as shown in Eq. (43).

$$2O_3 + C\ (\text{Organic}) \rightarrow CO_2 + 2O_2 \tag{43}$$

The above reaction is oxidation reaction, and the carbon dioxide and oxygen are oxidation products. The remaining amount of available ozone reacts with organic substances yielding various intermediate and end products. The reaction is termed ozonolysis reaction, and the end products are termed ozonolysis products.

In ozonolysis reactions, ozone reacts readily with unsaturated organic compounds, ($=C=C=$ or $-C\equiv C-$), adding all its three oxygen atoms at a double bond or a triple bond. The resulting compounds are termed ozonides. Decomposition of ozonides generally results in a rupture at the position of the unsaturated bond, causing the formation of simpler organic substances, such as ketones, aldehydes, acids, and so on.

A typical reaction of ozone is its addition to the carbon–carbon bond of an ethylenic compound (68). This is illustrated by Eq. (44) in Fig. 15.

The resulting ozone-olefin compound is known as an ozonide, as discussed earlier. Decomposition of the ozonide gives a mixture of oxygenated products containing carbonyl compounds. This is shown by Eq. (45) in Fig. 15. Ozonides are not isolated as such, because of their unstable and explosive nature, but are employed for the production of other chemical compounds. Useful products in good yields are obtained when the ozonides are reduced to produce carbonyl compounds or oxidized to produce carboxylic acids.

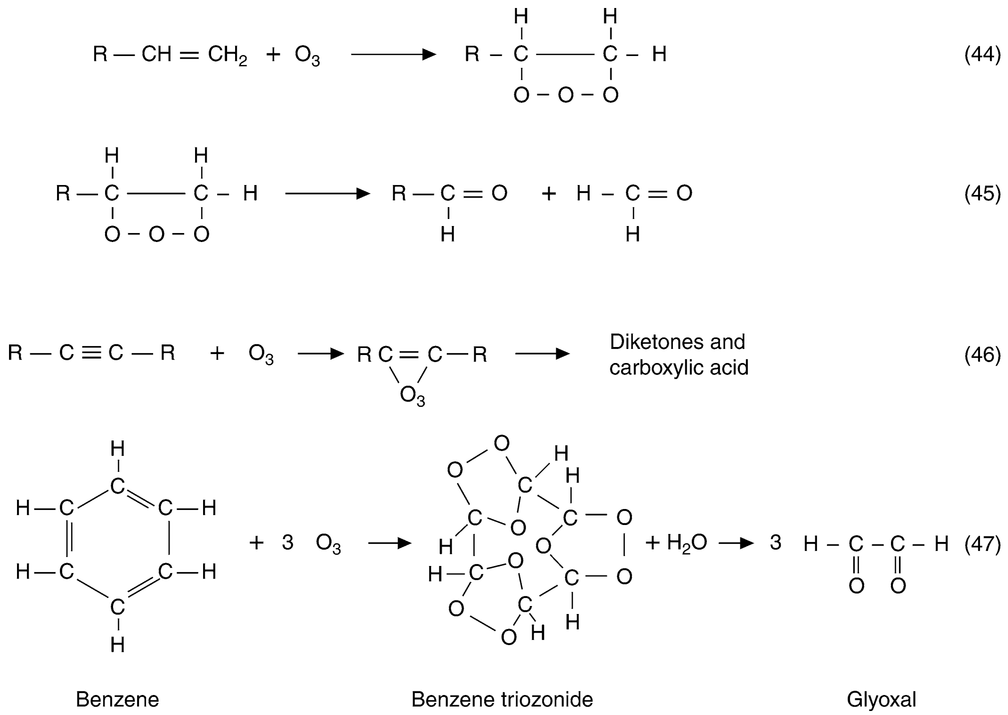


Fig. 15. Ozone reactions with organics-equations (44)–(47) (68).

Ozone also adds to the carbon–carbon triple bond of acetylenic compounds, as shown in Eq. (46) of Fig. 15. Very few detailed studies of the reaction have been made. The usual products isolated are diketones and carboxylic acids. Ozone adds very slowly to benzene. Each benzene molecule can add three molecules of ozone and yield three moles of glyoxal, Eq. (47) in Fig. 15. Methyl-substituted benzenes react more rapidly. Hexamethyl benzene reacts several thousand times more rapidly than benzene. Hydrolysis of the ozonide products produce the carbonyl compounds usually obtained in ozonolysis reactions. Equation (48) in Fig. 16 illustrates how 2 mol of methyl-substituted benzene react with 6 mol of ozone yielding 3 mol of glyoxal, 1 mol of diacetyl, and 2 mol of z-methylglyoxal (31,68).

In polynuclear aromatic compounds the various carbon bonds and atoms have different reactivities. The reaction with ozone is more complex and the composition of the products is difficult to predict. Oxidation at a carbon atom may, at times, predominate over ozone addition to a carbon–carbon double bond. While some aromatic compounds might add ozone rapidly and form ozonides, others may be oxidized to give quinones. Frequently, aromatic compounds react in both ways and the reaction mixture contains both oxidation and ozonolysis products.

Saturated hydrocarbons react very slowly with ozone at room temperature. However, at elevated temperatures, the reaction proceeds quite rapidly. Peroxides, ketones or aldehydes, alcohols, and acids are formed as the reaction products. Ethers are oxidized by ozone at the carbon next to the ether oxygen. Therefore, esters are found among the

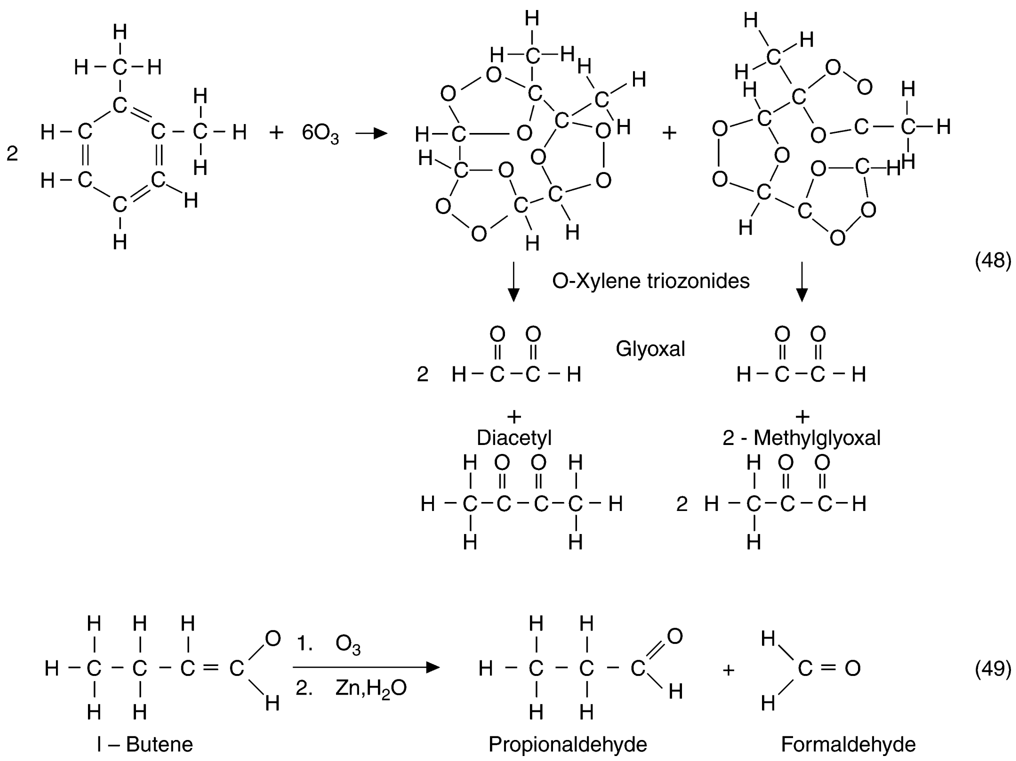


Fig. 16. Ozone reactions with organics-equations (48)–(49) (31,68).

oxidation products. Lactones are produced by the reaction of ozone and cyclic ethers. Ozonation of cyclic formals produce carbonates.

Organic sulfides are oxidized by ozone through sulfoxides (RSOR′), to sulfones (RSO_2R'). The intermediate sulfoxide may be isolated. Primary and secondary amines are only degraded by ozone but tertiary amines form tertiary amine oxides. Organic phosphates might be prepared by ozone oxidation of phosphites, and phosphine oxides are formed from phosphines. Besides the ozonolysis and oxidation reactions where stoichiometric amounts of ozone are reacted, there are catalytic reactions where experimental conditions determine the amount of ozone used. The preparation of peroxyacids from aldehydes is one of these reactions. Ozone is only a catalyst or initiator of the oxidation. In the production of cumene hydroperoxide intermittent ozonation of cumene is used.

Ozonolysis followed by hydrolysis is reliable, though at times dangerous as a result of the unstable and explosive nature of ozonides, which search for the double bond in organic compounds. This ozonolysis/hydrolysis reaction sequence can be illustrated by the difference in reaction products obtained by the reaction of ozone with 1-butene and 2-butene, as shown by Eq. (49) in Fig. 16 and Eq. (50a) in Fig. 17. Usually ozonized oxygen, containing up to 15% ozone, is passed at room temperature into a solution of the unsaturated compound. The ozonide formed is hydrolyzed either in the presence of a reducing agent to obtain aldehydes and ketones, or with hydrogen peroxide in acetic acid solution to get carboxylic acids (31).

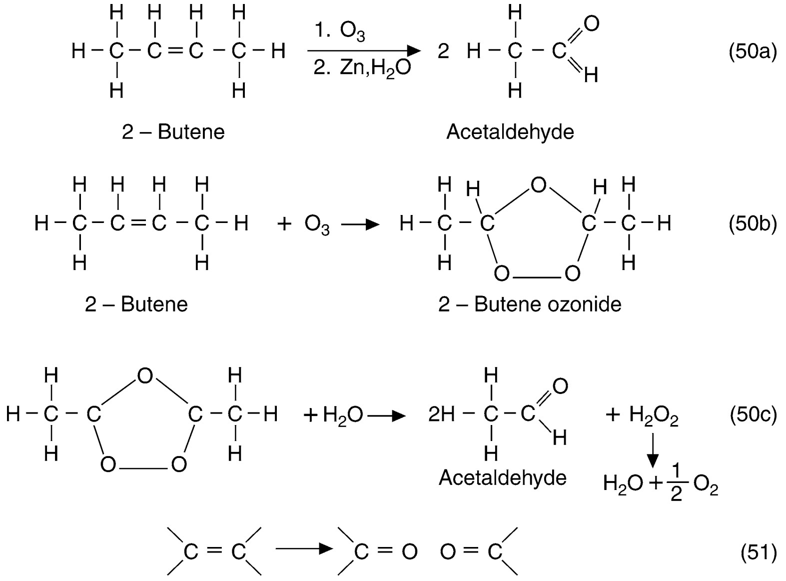


Fig. 17. Ozone reactions with organics-equations (50), (51) (31).

The decomposition of an ozonide can also be effected by catalytic hydrogenation. The over-all reaction is a breaking of the double bond with two carbonyl groups appearing in its place as represented by Eq. (51) in Fig. 17. Ozonides, like most substances with the peroxide (O–O) bonds, are very unstable and may explode violently and unpredictably. Ozonation must, therefore, be carried out with due care and caution. The ozonides are usually not isolated but are destroyed by hydrolysis with water to yield carbonyl compounds that are generally quite easy to isolate and identify. The ozonation of 2-butene followed by hydrolysis to form acetaldehyde is an example of this, as shown by Eqs. (50b,c) in Fig. 17.

Very few detailed studies have been made about the chemical pathways involved in ozonation of organic substances in water. In most studies in aqueous media, a co-solvent, a water emulsion, or a suspension has been employed. Pryde (69) has ozonized methyl oleate and methyl lineoleate in water media and decomposed the peroxidic ozonolyses products to aldehydes and/or carboxylic acids under various conditions. Criegee and Lohaus (70) has synthesized a peroxide from ozonolysis of cyclic sulfone in the presence of water.

Sturrock (71) has ozonized the aliphatic 9, 10 bond of phenanthrene in an aqueous *t*-butyl alcohol medium and obtained a dialdehyde. Ozonolysis of acetylenic compounds (C=C) in aqueous media have not been studied but presumably peroxides would be produced that readily decompose to carboxylic acids.

Ozonation of aromatic compounds appears to involve both 1, 3-dipolar cycloaddition at a carbon–carbon bond to give ozonolysis products, and electrophilic ozone attack at individual carbon atoms (68). Regarding the ease of ozone attack, the unsubstituted benzene ring is lesser reactive than is an olefinic double bond (68). Polycyclic aromatics such as phenanthrene, anthracene, and naphthalene fall in between in reactivity (68). Alkyl and other substituents, which activate for electrophilic attack facilitate ozone attack while those which deactivate for electrophilic attack drastically decrease the rate of ozone attack on the aromatic nucleus (68). However, benzene (72) and its homologs (68) react to give the ozonolysis products as shown by Eq. (47) Fig. 15 and Eq. (48) Fig. 16.

Considerable study has been given to the removal of phenolic waste from water (73). Phenols are more reactive with ozone than are most aromatics, and phenol itself has been oxidized to carbon dioxide, formic acid, glyoxal, and oxalic acid (74). Eisenhauer (73,75) has carried out a detailed study of the reaction of ozone with phenol in water solutions and has identified catechol and *O*-quinone as intermediary products. However, he did not establish that the reaction went solely through these intermediates. Muconic acid was assumed to be the next intermediate, followed by further ozonolysis of this unsaturated substance (73). Ozonation of other phenols and napththols have also been shown to occur readily (76–79).

The ozonation of polycylic hydrocarbons has been studied extensively (68,80,81), although most of the work has been done in nonaqueous media. Most aromatic and aliphatic unsaturated heterocycles readily react with ozone (68). An exception is the pyridine ring, which reacts very slowly. Various organic substances with a nucleophilic atom in their structure are readily oxidized by ozone (68).

Various amino acids and proteins have been ozonized in water solution, but the ozone attack appears to have occurred at sulfur (cystine), aromatic, or heterocyclic unsaturated carbon–carbon bonds rather than nitrogen, (82). When reactive groupings are not present in organic molecules, ozone attack on carbon–hydrogen bonds becomes possible. Such reaction occurs readily with aldehydes (83–85), ethers (86,87), alcohols (88), and hydrocarbons (89,90) or hydrocarbon groupings (91) having secondary or tertiary carbon–hydrogen bonds. Through these reactions, aldhydes are converted to carboxylic acids, primary and secondary alcohols to carborylic acids, aldehydes or ketones, ethers to alcohols and esters, and hydrocarbons to alcohols and ketones. There appears to be general agreement that these ozonation reactions involve a hydrotrioxide intermediate. However, the exact mode of formation of the intermediate and how it is converted to products is not clear.

The ozonation of ethers has recently been carried over to acetals, including sugar glycosides, to yield esters (92). Evidence is also available about the ozonation of sugar alcohols and of polysaccharides in aqueous medium (93), including the degradation of wood pulp (94–96). A simple carbon–hydrogen bond ozonation, which has been performed in water solution is that of malonic acid (97). The methylene group (CH_2) was converted to an alcohol and a ketone function. Oxalic acid and carbon dioxide were also produced.

Because ozonides are active oxygen compounds, they could be used as oxidizing agents, polymerization catalysts, bleaching agents, and germicides. But their instability makes it difficult to prepare them in good yields and to use them safely in reactions. Ozonides or ozonolysis products have at times exploded on standing. Commercially, ozonides are not handled as such, but serve only as intermediates to more stable products. The reduction of some organic ozonides will produce aldehydes in good yields and oxidation will produce carboxylic acids. Thermal composition of an ozonide will produce a mixture of aldehydes, acids, peroxy compounds, and some other byproducts. The nature of solvents determines what ozonolysis products will be obtained. In nonpolar hydrocarbon solvents, ozonides will be formed. In more polar solvents, a mixture of ozonides, peroxides, aldehydes, and acids will be produced.

Ozonolysis products are thermally unstable. The reaction must be maintained within a specific temperature range in order to prepare these compounds. Because the ozone addition reaction is highly exothermic, reaction vessels must be cooled in order to maintain the desired reaction temperature.

Amines and amino acids may be prepared by reductive amination of oxonides or ozonolysis products. Ozone adds three oxygen atoms to the double bond. Therefore, further oxidation of the ozonide is required to give two moles of acids. Oxidation with air or oxygen will give satisfactory results in most cases. However, some ozonides may be more difficult to oxidize, and in these cases, oxidation with peroxyacids is required.

Humic acid is a major precursor for trihalomethane formation. Glaze et al. (98) have reported on the ability of ozone to destroy humic acid. Guirguis et al. (99) confirmed that ozone makes organic compounds better adsorbed by activated carbon.

Prengle et al. (100) reported that ozone plus UV light could reduce an organic pesticide, malathion, to carbon dioxide and water and simultaneously could reduce the total organic carbon in water. Likewise, Richard (101) found that ozone could degrade two other pesticides, parathion and marathion, to simple phosphoric acid.

8. OXYGENATION AND OZONATION SYSTEMS

8.1. Oxygenation Systems

The use of pure oxygen or oxygen-enriched air instead of air alone can improve upon many conventional water and waste treatment technologies. Some of the developments in this regard include the following:

a. Oxygenated-activated sludge process for wastewater treatment and/or nitrification.
b. Autothermal thermophilic oxygen digestion process for sludge treatment.
c. DO flotation process for iron and manganese removal from water.
d. Oxygenation–ozonation process for sludge management and wastewater reclamation.
e. Improved ozone generation technology using pure oxygen or oxygen-enriched air.

Potential sources of supply for oxygen include on-site oxygen generation plants, transport of liquid oxygen to the site, and use of an oxygen gaseous pipeline supply. On-site oxygen generation systems are the most economical and desirable form of oxygen supply for the aforementioned five applications, provided the plant is large enough (i.e., at least 1 MGD) to economically handle the capital cost for such installation.

A liquid oxygen supply system is useful for small plant's operation, or as back-up oxygen supply and peak load equalization for plants of any size. Although the unit oxygen cost is generally high, the small plants could find this to be a good option because of the saving of cost for capital investment. A pipeline oxygen supply system is practical if the treatment plant is in the vicinity of an oxygen generation facility.

There are two basic oxygen generation systems which are as follows:

a. The traditional cryogenic air separation (CAS) process for the large size plants.
b. The pressure-swing adsorption (PSA) process for the somewhat smaller and more common plant sizes.

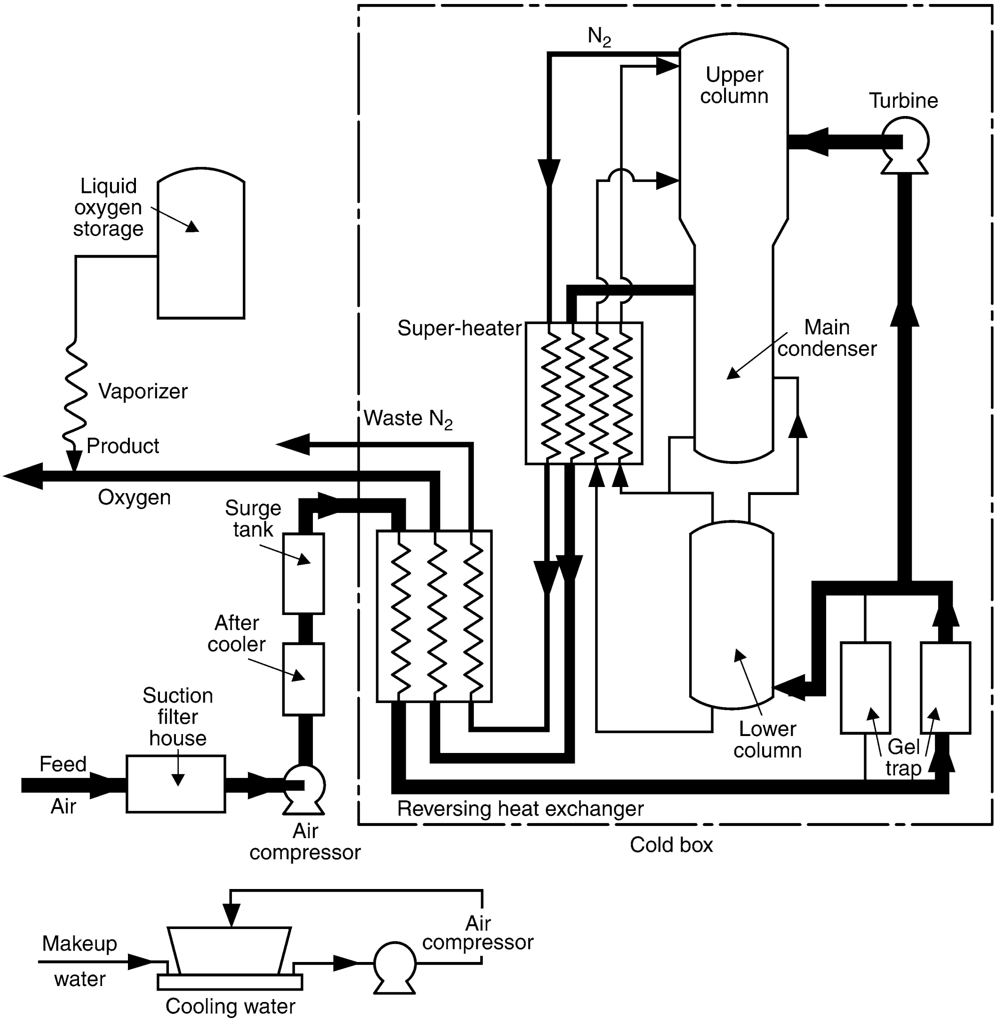


Fig. 18. CAS system for oxygen production (*Source:* Union Carbide Corp.).

The standard CAS process (Fig. 18) involves the liquefaction of air followed by fractional distillation to separate it into its components (mainly nitrogen and oxygen). The entering air is first filtered and compressed. It is then fed to the reversing heat exchangers, which perform the dual function of cooling and removing the water vapor and carbon dioxide by freezing these mixtures out onto the exchanger surfaces. This process is accomplished by periodically switching or "reversing" the feed air and the waste nitrogen streams through identical passes of the exchangers to regenerate their water vapor and carbon dioxide removal capacity. The air is next processed through "cold end gel traps" which are adsorbent beds, which remove the final traces of carbon dioxide as well as most of the hydrocarbons from the feed air. It is then divided into two streams, one of which feeds directly to the "lower column" of the distillation unit and the other is returned to the reversing heat exchangers and partially warmed to provide the required temperature difference across the exchanger. This second stream is then passed through

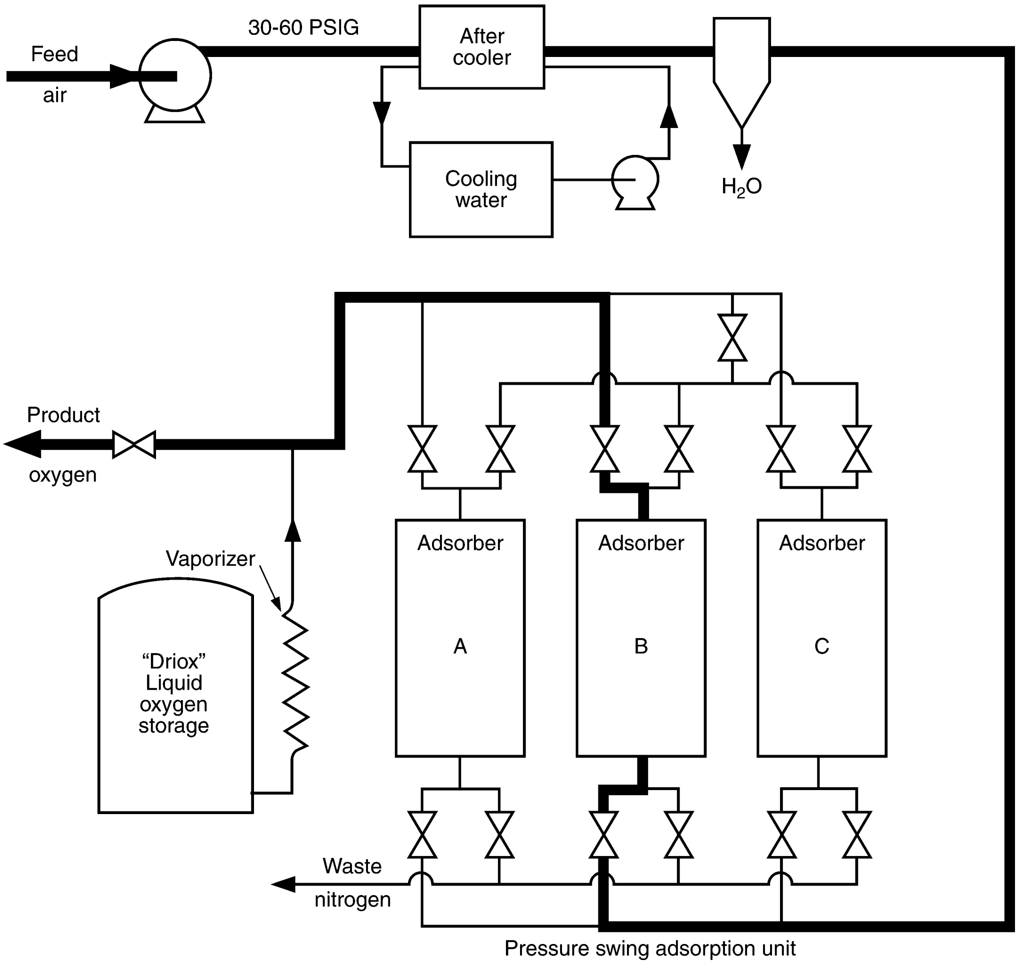


Fig. 19. Pressure swing adsorption system for oxygen production (*Source:* Union Carbide Corp.).

an expansion turbine and fed into the "upper column" of the distillation unit. An oxygen-rich liquid exits from the bottom of the "lower column" and the liquid nitrogen from the top. Both streams are then sub cooled and transferred to the upper column. In this column the descending liquid phase becomes progressively richer in oxygen until what collects in the condenser reboiler is the oxygen product. This oxygen stream is continually recirculated through an adsorption trap to remove all possible residual traces of hydrocarbons. The nitrogen exits from the top portion of the "upper column" and its heat is exchanged with the oxygen product to recover all available refrigeration and to regenerate the reversing heat exchangers as discussed earlier.

The pressure swing adsorption system uses two (or more) adsorbent vessels to provide a continuous and constant flow of oxygen gas. Figure 19 shows a PSA system with three adsorbers. In operation, the feed air is compressed by a nonlubricated compressor. This compressed air is separated into oxygen and nitrogen rich streams as it flows through one of the adsorbent vessels. The adsorbent is a granular material (molecular sieve), which attracts and traps (adsorbs) the carbon dioxide, water, and

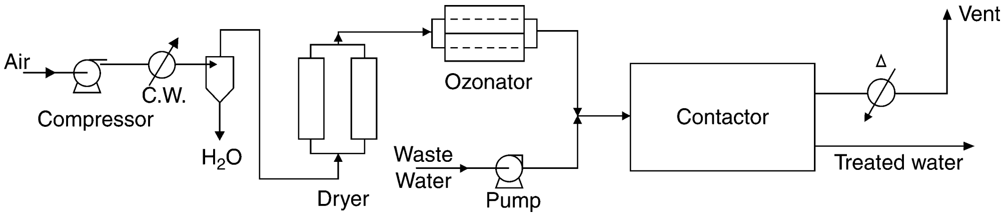


Fig. 20. Open-loop ozonation system using air (*Source*: W. R. Grace & Co.).

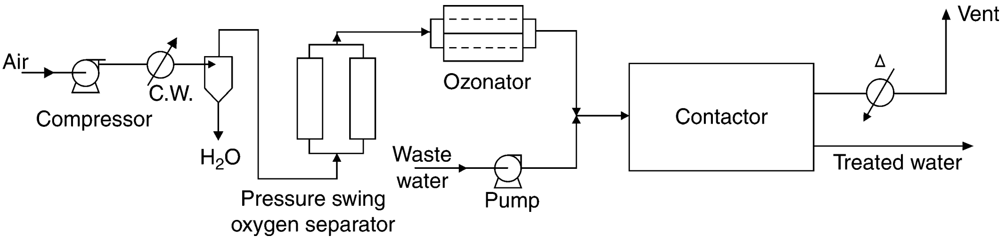


Fig. 21. Open-loop ozonation system using oxygen-rich air (*Source:* W. R. Grace & Co.).

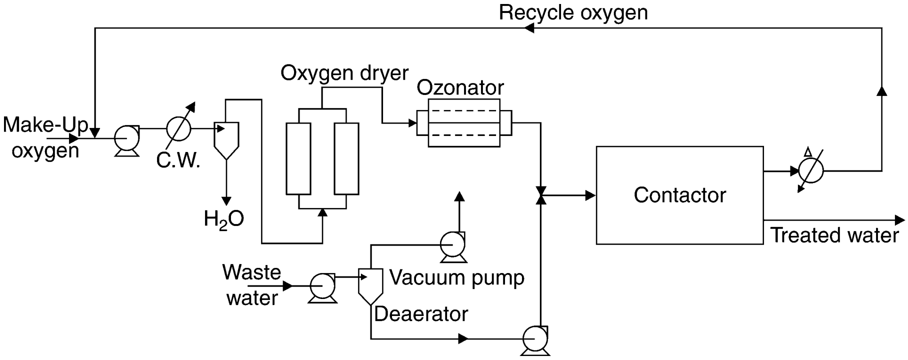


Fig. 22. Closed-loop ozonation system using oxygen (*Source:* W. R. Grace & Co.).

nitrogen gas, producing a relatively high purity oxygen product. While one bed is adsorbing, the other beds are in various stages of regeneration. The PSA oxygen generator operates on a pressure swing, an adsorption concept in which the oxygen is separated from the feed air by adsorption at high pressure and the adsorbent is regenerated by blow down to low pressure. The process operates on a repeated cycle that has two basic steps, adsorption and regeneration. During the adsorption step feed airflow through one of the adsorbent vessels until the adsorbent is partially loaded with impurity. At that time the feed airflow is switched to another adsorber and the first adsorber is regenerated. During the regeneration step the impurities (carbon dioxide, water, and nitrogen) are cleaned from the adsorbent so that the cycle (adsorption–generation) can be repeated.

Regeneration of the adsorber is carried out in three basic steps which are as follows:

a. The adsorber is depressurized to atmospheric pressure to remove some impurities from the adsorbent and to make it easy to remove the remaining impurities.

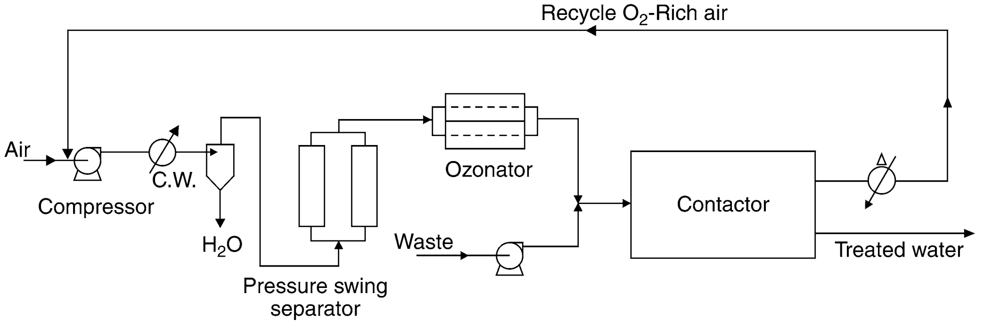


Fig. 23. Closed-loop ozonation system using oxygen-rich air (*Source:* W. R. Grace & Co.).

Table 20
Removal data of Conventional Pollutants by Ozonation

Pollutant/parameter	Concentration Influent	Effluent	Removal %
Classical pollutants (mg/L)			
BOD_5	5800	5200	10
COD	77,000	12,000	84
TSS	64	140	NM
Oil and grease	130	4	97
Total phenol	47	0.13	>99
Toxic pollutants (μg/L)			
Cyanide	560	1500	NM
Zinc	2200	90	96

Source: US EPA.
Blanks indicate data not available.
NM, not meaningful.
Sampling: equal volume grab samples collected throughout an 8 h/d; average of 2 d sampling.

b. The adsorber is purged with product oxygen to clean the remaining impurities.
c. The adsorber is repressurized to adsorption pressure and is again ready to separate the feed air.

8.2. Ozonation Systems

Generally there are four basic ozonation systems used in water and waste treatment. Figures 20–23 illustrate the four systems. When the least cost basic system has been arrived at, there remains only the contactor that must be designed to finalize the ozone treatment system. For details and more information on ozonation technology, monitoring, and control instrumentation the reader is referred to refs. (102–106).

In the first system compressed air is cooled to remove moisture and is fed to a dryer. The dry air is fed to the ozone generator and the ozone-air solution (1–3% ozone by weight) is mixed with the incoming water or waste in a contactor. The treated effluent and gases leave the contactor separately. Any excess ozone in the effluent soon decomposes to oxygen while the ozone in the waste gas should be destroyed by heat or by chemical or catalytic decomposition. This system is applicable to very small installations.

Table 21
Removal of Toxic Pollutants by Ozonation

Pollutant	Removal (%)
Heavy metals	
Antimony	76
Arsenic	45
Xylene	97
Cadmium	98
Chromium	52
Copper	75
Lead	98
Mercury	75
Nickel	73
Silver	45
Organic chemicals	
Bis (2-ethylhexyl) phthalate	72
Butyl benzyl phthalate	99
Carbon tetrachloride	75
Chloroform	58
Dichlorobromomethane	99
Diethyl Di-*N*-butyl phthalate	97
Diethyl phthalate	99
Di-*N*-octyl phthalate	78
N-nitrosodiphenylamine	66
2,4-Dimethylphenol	99
Pentachlorophenol	19
Phenol	57
Dichlorobenzene	76
Ethylbenzene	65
Toluene	39
Naphthalene	77
Anthracene/phenanthrene	81

Source: US EPA.

The second system is similar to the first but is somewhat more cost efficient for larger installations because air has been replaced by oxygen-enriched feed stream. The use of a pressure swing adsorption technique for producing oxygen-enriched air will reduce the capital and operating costs of the ozone generator.

The third system for ozone treatment is similar to the first but oxygen is fed to the generator and oxygen rich off-gas is recycled to the front end of the loop. There is an additional de-aeration step to remove nitrogen from the wastewater before its treatment so that it does not build up in the gas recycle.

In the fourth system, air is enriched to about 40% oxygen in the startup cycle. In each successive cycle, the recycled gas is cleaned, dried, enriched in oxygen and any excess ozone decomposed by the catalytic effect of the molecular sieves used in the pressure swing separator. The system can be programmed so that the economic point of ozone generation vs oxygen generation (probably around 80% oxygen) is achieved. The other 20% will consist of nitrogen, carbon dioxide, and argon. By mixing adsorbents in the

Table 22
Removal data of Cyanide by Ozonation

Flow rate (L/min)	Ozone feed rate (g/h)	Cyanide: Concentration (μg/L) Influent	Cyanide: Concentration (μg/L) Effluent	Cyanide: Removal (%)
3200		900	<20	>97
9.5	3	360	20	94
9.5	3	160	18	89
4.9	6	200	95	51

Blanks indicate data not available.
Source: US EPA.
Sampling: Grab

pressure swing adsorption separator, nitrogen, CO_2, and water will be removed in each cycle and oxygen increased. Air of course will be added for makeup. At less than the 10% level, CO_2 and argon have little effect on ozone production efficiency and nitrogen has a positive effect. This system provides the following functions with a single pressure swing adsorption separator in a closed loop with the ozone generator and contact system.

a. Oxygen enrichment.
b. Removal of nitrogen and CO_2.
c. Drying to −100°F dew point.
d. Catalytic decomposition of excess ozone.

8.3. Removal of Pollutants from Waste by Ozonation

The removal data compiled by the US Environmental Protection Agency (US EPA) are summarized in Tables 20, 21, and 22, The tables clearly show that the ozonation process can efficiently remove not only the classical pollutants such as BOD, COD, TSS, oil, grease, and phenol (Table 20), but also heavy metals and toxic organics (Tables 21 and 22). More information on ozonation and pressurized ozonation processes can be found elsewhere (107–111). Wang also reported the UV is an excellent deozonation process for removal of residual ozone from a treated waste stream (112).

NOMENCLATURE

B	Concentration of ozone in liquid/Concentration of ozone in gas, reduced to STP The Bunsen coefficient of solubility
C	Residual concentration of disinfectant (mg/L)
H	Henry's constant, mole/atm (mole fraction of ozone in solution/partial pressure of ozone in gas, in atm)
k_1, k_2, and k_3	Rate constants
L	Lethality coefficient ($[\text{mg/L}]^{-1}\ \text{min}^{-1}$)
M	A chemical element
(OH^-)	Hydroxide concentration (g-mole/L)
T	Temperature (°K)
T	Time in minutes for 99% microorganism destruction (2-log destruction)

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9
Low-Temperature Thermal Treatment Processes

Lawrence K. Wang, Clint Williford, Wei-Yin Chen, and Nazih K. Shammas

CONTENTS

1. INTRODUCTION

There are two kinds of thermal processes for sludge treatment: (a) heat conditioning: a conditioning process which prepares sludge (i.e., mainly biosolids) for dewatering on filter presses or vacuum filters without the use of chemicals and (b) heat-drying: a process which evaporates water from sludge by thermal means (1–11). Ambient air-drying processes are dealt with in another two chapters: "Evaporation processes" and "Drying beds."

2. HEAT CONDITIONING PROCESS

2.1. Process Description

The heat conditioning process (Fig. 1) involves heating sludge (i.e., biosolids) to temperatures of 140–240°C (285−460°F), for a period of 15–60 min under pressures of 250−400 lb/in.^{2}g (1720–2760 kN/m^2). It is essentially a thermal, nonchemical conditioning process, which prepares sludge for dewatering in vacuum filters or filter presses without the use of chemicals. In addition, the sludge is sterilized, generally stabilized, and rendered inoffensive. Heat results in coagulation of solids, a breakdown in the cell structure of sludge, and a reduction of the water affinity of sludge solids.

Several proprietary variations exist for thermal treatment by heat conditioning. In these systems, sludge is passed through a heat exchanger into a reactor vessel, where steam is injected directly into the sludge to bring the temperature and pressure into the necessary ranges. In one variation, air is also injected into the reactor vessel with the sludge. The detention time in the reactor is approx 30 min. After heat conditioning, the sludge passes

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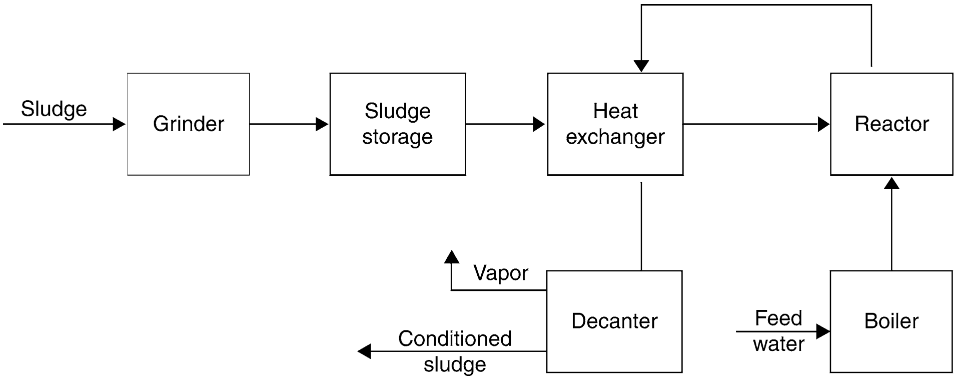


Fig. 1. Heat conditioning process flow diagram. Source: US EPA (3).

back through the heat exchanger to recover heat, and then is discharged to a thickener-decant tank. The thickened sludge may be dewatered by filtration or centrifugation to bring its solids content up to 30–50%. The sludge may be ground prior to heat conditioning.

2.2. Process Applications and Limitations

2.2.1. Applications

The process of heat-treating sludge was first introduced in 1935, but has only become common during the last decade. About 400 units are currently in operation in the United States. Heat-conditioning is practised as a sludge conditioning method to reduce the costs of sludge dewatering and ultimate disposal. The benefits of heat conditioning as follows:

a. Improved dewatering characteristics of treated sludge without chemical conditioning.
b. Innocuous and sterilized sludge is obtained, which is suitable for ultimate disposal by a variety of methods including land application.
c. Less nuisance value.
d. Suitable for the types of sludge which cannot be stabilized biologically.
e. Insensitive to variations in sludge composition.
f. Reduction in subsequent incineration energy requirements.
g. Reduction in size of subsequent vacuum filters and incinerators.

2.2.2. Limitations

The process has very high capital and operating costs, and may not be economical at small treatment plants. Specialized supervision and maintenance is required owing to the heat and pressures involved. Expenditure on expensive material is necessary to prevent corrosion and withstand the operating conditions. Heavy metal concentration in sludges are not reduced by heat conditioning, and further treatment of sludges with high metallic concentrations may be required if the sludge is to be applied to cropland. The sludge supernatant and filtrate recycle liquor are strongly colored and contain very high concentrations of soluble organic compounds and ammonia nitrogen in addition to the need for collecting an treating the odorous gas stream, in some cases the liquid sidestreams must also be pretreated before its return to the head of the treatment plant.

2.3. Design Considerations

2.3.1. Target Process Performance

Heat conditioning is a process intended to enhance the performance of subsequent operations. In the process itself pathogens are destroyed and 30–40% of the volatile suspended solids are solubilized. Dewatering performance can be increased to a solids capture of over 95% and a sludge solids content of up to 50%. Chemicals for dewatering are not normally required. Corrosion control aids may be required for the boiler and/or the process.

The liquid sidestream (recycle liquor) produced represents about 50% of sludge flow (by volume). Quality wise, the liquor composition varies greatly with reported concentrations of biochemical oxygen demand (5000–15,000 mg/L), chemical oxygen demand (10,000–30,000 mg/L), NH_3–N (30–800 mg/L), P (140–250 mg/L), total suspended solids (9000–12,000 mg/L), volatile suspended solids (8000–10,000 mg/L), and pH 4.0–6.0. This stream is generally amenable to biological treatment but it can contribute as much as 30–50% of the original loading to a wastewater treatment plant. If the plant has not been designed for this additional load, pretreatment prior to return may be necessary. Some noncondensable gases may be generated which will require combustion or disposal.

2.3.2. Design Considerations

Complete heat treatment systems are generally proprietary, and manufacturers supply most of the common systems. The major equipments common to these processes are grinders, sludge feed pumps and handling equipment, heat exchangers, reactors, boilers, and separators.

The following are the design criteria for a heat conditioning process:

a. Temperature (140–240°C).
b. Pressure (250–400 lb/in.^{2}g).
c. Detention time (15–60 min).
d. Steam consumption is about 600 lb/1000 gal of sludge.

2.3.3. Environmental Impact and Operational Considerations

Recycled liquor sent to the head of a plant can cause plant upsets owing to very high organic loadings. The process can result in offensive odor production if proper odor control is not practiced. A colored effluent may also result, requiring additional processing if discharge standards prohibit this condition.

The composition of the recycle liquor can vary among the various processes. Some liquors may contain a high proportion of nonbiodegradable matter. This matter is largely humic acids, which results in an unpleasant odor and taste if present in water, which has been chlorinated prior to its use for domestic supply. If industrial waste of various types are included in the wastewater to be treated, the actual chemical composition of the liquor resulting from heat treatment of the sludge should be determined by a detailed chemical analysis. A possible treatment process for highly polluted liquor could consist of filtration, aeration, and activated carbon adsorption for nonbiodegradable organics.

Table 1
Fuel Requirements of Sludge Treatment by Heat Conditioning Process

Treatment capacity (gal/min)	Air supply conditions	Fuel requirements (M Btu/yr)
1	Without air	280
10	Without air	2750
100	Without air	26,000
1	With air	420
10	With air	4150
100	With air	41,000
1	Maximum air	500
10	Maximum air	5000
100	Maximum air	49,000

Source: US EPA (3,21).

Table 2
Electrical Energy Requirements of Sludge Treatment by Heat Conditioning Process

Treatment capacity (gal/min)	Air supply conditions	Electrical energy (kWh/yr)
1	No air	30,000
10	No air	70,000
100	No air	425,000
600	No air	2,800,000
1	With air	36,000
10	With air	130,000
100	With air	600,000
600	With air	6,000,000

Source: US EPA (3,21).

2.3.4. Energy and Costs

Tables 1 and 2 show the energy consumption of the heat conditioning process, which is estimated based on the following assumptions:

a. Reactor conditions (300 lb/in.^{2}g at 350°F).
b. Heat exchangers ($T = 50$°F).
c. Continuous operation.
d. Electricals include all pumping, grinding, air compressing, and post-thickening drives.
e. Fuel is used to produce steam to bring reactor to operating temperature.

Tables 3 and 4 show the construction cost and operation and maintenance costs, respectively, of the heat conditioning process at various wastewater flows.

Construction costs include: sludge feed pumps, grinders, heat exchangers, reactors, boilers, gas separators, and buildings. Cost is related to average wastewater flow by the following: sludge quality is about 1900 lb/MG (undigested, combined thickened

Table 3
Construction Costs of Heat Conditioning Process Facilities

Wastewater flow of WWTP (MGD)	Construction cost of facility[a] (million USD)
0.1	0.258
1	1.162
10	2.582
100	10.844

Source: US EPA (3,21).
[a] Note: January 2002 cost data (ENR CC Index = 6390.21).

Table 4
Operation and Maintenance Costs of Heat Conditioning Process Facilities

Wastewater flow of WWTP (MGD)	Cost elements	Annual O&M costs[a] (million USD)
0.1	Power	0.00194
1	Power	0.00568
10	Power	0.02582
100	Power	0.18074
0.1	Materials	0.01549
1	Materials	0.01807
10	Materials	0.03098
100	Materials	0.10586
0.1	Fuel	0.00070
1	Fuel	0.00697
10	Fuel	0.06971
100	Fuel	0.69714
0.1	Labor	0.10328
1	Labor	0.11619
10	Labor	0.18074
100	Labor	0.54222
0.1	Total	0.12141
1	Total	0.14717
10	Total	0.30726
100	Total	1.52596

Source: US EPA (3,21).
[a] January 2002 cost data.

primary plus secondary sludge); solids concentration is about 4.5%, sludge flow is about 3.8 gal/min/MGD based on 8000 operating h/yr. Fuel cost is for steam generation. Engineering News Record (ENR) Index is about 6390.21.

To adjust costs for design factors different from the earlier mentioned, enter tables at effective flow (Q_E).

$$Q_E = Q_{\text{design}} \times (\text{New design sludge mass})/(1900 \text{ lb/MG}) \tag{1}$$

3. HEAT DRYING PROCESS

3.1. Process Description

Conventional heat drying is usually preceded by mechanical dewatering and may be followed by air pollution control devices and systems that alter the form of the dried material. Mechanical dewatering is discussed in detail in another chapter. It is an important pretreatment step because it reduces the volume of water that must be removed in the drying process. In the dryer, water that has not been mechanically separated is evaporated without decomposing the organic matter in the sludge solids. This means that the solids temperature must be kept between 140 and 200°F (60 and 93°C). A large portion of the dried sludge is often blended with the sludge fed to the dryer, making the drying operation more efficient by reducing agglomeration (large balls of sludge), thus exposing more solids surface area to the drying medium. Dried sludge and exhaust gases are separated in the dryer itself and/or in a cyclone. The gas stream can be passed through a pollution control system for removal of odors and particulates. The dried sludge is then sent to a finishing step such as pelletizing or bagging, or it is stored in bulk for marketing or use in the next portion of the sludge management scheme.

Sludge is heat dried at temperatures too low to destroy organic matter. Air is used to carry away the water vapor. The designer specifies the actual conditions of drying, which include temperature, humidity, detention time, velocity, and direction of flow of the gas stream across the drying surface. There are three well-defined stages in heat drying: initial drying, steady-state drying, and final drying (2).

3.1.1. Initial Drying

During this stage, the sludge temperature and the drying rates are increased to equal the steady-state conditions of the second stage. Stage one is usually short; little drying occurs during this time.

3.1.2. Steady-State Drying

In comparison with other stages, the sludge is processed in this stage for the longest time. The surfaces of the sludge particles are completely saturated with water. Surface water is replaced with water from the interior of the solid as soon as it is evaporated. Drying proceeds similar to how water evaporates from a pool. The solid does not significantly influence the drying rate. For this drying period, the temperature at the sludge/gas interface is ordinarily kept at the wet-bulb temperature of the gas. As long as unbound surface moisture is present, the solid is heated only to the wet-bulb temperature of the gas; solids may therefore be dried with fairly hot gases and not attain elevated temperatures. For example, the wet-bulb temperature is 133°F (56°C) for a gas stream that has an absolute humidity of 0.01 lb water per pound dry air and a temperature of 600°F (316°C).

3.1.3. Final Drying

The final stage occurs when sufficient water has evaporated and the solid surface is only partially saturated. Surface water is evaporated more rapidly than it can be replaced by water from the interior of the solid. As a consequence, overall drying rates are markedly lower in stage 3 than in stage 2. During this period, the temperature of the solid/gas interface increases because latent heat cannot be transferred from the sludge to the gas phase as rapidly as sensible heat is received from the heating medium.

3.2. Design Considerations

3.2.1. Sludge Moisture Content

Sludge moisture content is normally expressed in moisture (%), solids (%), or pounds water per pound dry sludge. The minimum sludge moisture content practically attainable with heat drying depends upon the design and operation of the dryer, moisture content of the sludge feed, and the chemical composition of the sludge. For ordinary domestic wastewater sludges, sludge moisture contents as low as 5% may be achieved. Chemical bonding of water within the sludge, which can occur through chemical addition for sludge conditioning, or chemicals present in industrial sludges can increase the amount of water retained in the dried products beyond the 5% moisture level.

3.2.2. Humidity and Mass Transfer

Humidity is the measure of the moisture content of the gas phase at a given temperature and is important to consider when determining drying rates. Absolute humidity is the measure of the weight of water per unit weight of dry gas (e.g., pounds water per pound dry air). In heat drying of sludge, water is transferred to the gas phase. The driving force for transfer is the difference between absolute humidity at the wet solid/gas interface and the absolute humidity in the gas phase. The transfer rate (i.e., the drying rate) can be described by the following equation:

$$W = K_y A\,(Y_s - Y_a) \tag{2}$$

where W is the rate of drying (lb water/h [kg/h]); K_y is the mass transfer coefficient of the gas phase (lb water/h/ft^2/unit of humidity difference [kg/h/m^2/unit of humidity difference]); A is the area of wetted surface exposed to drying medium (ft^2 [m^2]); Y_s is the humidity at the sludge/gas interface temperature (lb water/lb dry gas [kg/kg]); and Y_a is the humidity of the gas phase (lb water/lb dry gas [kg/kg]).

3.2.3. Temperature and Heat Transfer

In heat drying, the difference in temperature between the heating medium and the sludge/gas interface provides the driving force for heat transfer. Dryers are commonly classified on the basis of the predominant method of transferring heat to the wet solids being dried. These methods include convection, conduction, and radiation (2).

3.2.3.1. CONVECTION (DIRECT DRYING)

Heat transfer is accomplished by direct contact between the wet sludge and hot gases. The sensible heat of the inlet gas provides the latent heat required for evaporating the water. The hot gases carry off the vaporized liquid. Direct dryers are the most common type used in heat drying of sludge. Flash dryers, direct rotary dryers, and fluid bed dryers employ this method. Convective heat transfer is described by Eq. (3).

$$q_{\text{conv}} = h_c A\,(t_g - t_s) \tag{3}$$

where q_{conv} is the convective heat transfer (Btu/h [kJ/h]); h_c is the convective heat transfer coefficient (Btu/h/ft^2/°F [kJ/h/m^2/°C]); A is the area of wetted surface exposed to gas (ft^2 [m^2]); t_g is the gas temperature (°F [°C]); and t_s is the temperature at sludge/gas interface (°F [°C]).

3.2.3.2. CONDUCTION (INDIRECT DRYING)

Heat transfer is accomplished by contact of the wet solids with hot surfaces. The vaporized liquid is removed independently from the heating medium. The thin film dryer employs this principle. Conductive heat transfer is described by Eq. (4).

$$q_{\text{cond}} = h_{\text{cond}} A \,(t_m - t_s) \tag{4}$$

where q_{cond} is the conductive heat transfer (Btu/h [kJ/h]); h_{cond} is the conductive heat transfer coefficient (Btu/h/°F [kJ/h/°C]); A is the area of heat transfer surface (ft^2 [m^2]); t_m is the temperature of drying medium—for example, steam (°F [°C]); and t_s is the temperature of sludge at drying surface (°F [°C]).

The conductive heat transfer coefficient (h_{cond}) is a composite term that includes the effects of the heat transfer surface and sludge-side and medium-side films. Descriptions of methods for computing h_{cond} are available in handbooks, textbooks, and from dryer manufacturers (6–9).

3.2.3.3. RADIATION (INFRARED OR RADIANT HEAT-DRYING)

Heat transfer is accomplished by radiant energy supplied by electric resistance elements, by gas-heated incandescent refractories that also provide the advantage of convective heating, or by infrared lamps. The Shirco Company furnace and multiple-hearth furnaces are examples of drying equipment that use radiant heat. Radiation heat transfer is described by Eq. (5).

$$q_{\text{rad}} = \varepsilon_s A \sigma (t_r^{\,4} - t_s^{\,4}) \tag{5}$$

where q_{rad} is the radiation heat transfer (Btu/h [kJ/h]); ε_s is the emissivity of the drying surface (dimensionless); A is the sludge surface area exposed to radiant source (ft^2 [m^2]); σ is the Stefan–Boltzman constant (1.73×10^{-9} Btu/h/ft^2/°R) (4.88×10^{-8} kcal/m^2/h/K); t_r is the absolute temperature of the radiant source (°R [K]); and t_s is the absolute temperature of the sludge drying surface (°R [K]).

This discussion of heat drying is brief; the reader is referred elsewhere for more information (6–10). Equations for mass and heat transfer rates, and for associated drying times for specific dryer types are discussed in detail in these references. It is often difficult to determine appropriate values of mass and heat transfer coefficients to be used in these equations. Thus, results predicted by the equations and results obtained in design practice may be divergent, perhaps critically so. The most usable design information is obtained by testing with actual process feeds under conditions closely simulating prototype operation. Many dryer manufacturers provide such testing services.

3.2.4. Design Criteria

There are several common features of heat drying processes for which general design criteria can be developed.

3.2.4.1. DRYING CAPACITY

The number and size of the dryers depend on the type of drying operation contemplated. If the dryers are operated continuously, extra drying capacity is needed so that all sludge produced can be dried while maintenance and repair are being performed. In cases where noncontinuous operation (e.g., 40 h/wk) is envisioned or where only one dryer is installed, the dryer(s) must have sufficient evaporative capacity to handle all the sludge, including that generated when the dryers are not on-line. In the latter case, wet sludge storage requirements may be significant.

3.2.4.2. Storage Requirements

The design engineer should consider storage requirements for both the wet sludge feed and the dried product. Sufficient wet sludge storage should be provided to allow orderly shutdown of continuously operated drying processes (approx 3 d production at a peak rate). Storage for the dried product depends on the final disposal arrangement. Sales of the product are likely to be seasonal, and considerable storage may be necessary unless bulk buyers provide off-site storage. If the dried product is burned as a fuel or undergoes further processing, then the subsequent steps in the sludge-processing system will indicate the storage requirements. Dust can become a problem if the dried product is stored in bulk and is not pelletized. In some cases, the material should be appropriately contained.

3.2.4.3. Heat Source

The large amount of energy required for heat-drying dictates that close attention has to be given to the source used to heat the drying medium. Natural gas and fuel oil are most frequently used but are becoming more expensive, and shortages have occurred in the past few years. Energy recovery within the heat drying system itself provides one way of reducing energy usage; for example, heat exchangers can be used to recover heat from the exhaust gases. Another method is recovery of heat from a power source within the plant for example, Milwaukee recovers waste heat from gas turbine exhausts. The dried sludge itself has a fuel value and may be used as a heat source for the drying medium. Komax Systems, Inc. (12) provide process equipment for direct steam heating of sludges.

3.2.4.4. Airflow

Airflow is an important consideration in the design of direct dryers. Airflow may be cocurrent, countercurrent, or crossflow. In direct drying, cocurrent flow offers the advantage of higher thermal efficiency owing to rapid cooling of the heating medium near the feed end with concomitant reduced heat losses through the dryer structure. In addition, the dried sludge is not subjected to high-gas temperatures near the discharge end, as it would be in counterflow operation. This is advantageous because it minimizes distillation of odorous materials, and increases thermal efficiency to a certain extent somewhat by reducing heat loss with the dried sludge.

The rate of airflow is a function of the dryer design. However, turbulent conditions must be maintained to ensure intimate contact between the warm air and wet sludge. Dusting problems may limit airflow rate. Huber Technology Inc. (13) has developed a low temperature dryer, which is also called recirculation air sludge dryer, using recirculation of airflow for low temperature sludge heating and drying.

3.2.5. Environmental Impact and Operational Considerations

Heat drying of sludge produces a material that usually contains 10% or less moisture, a moist gas stream that is ejected to the atmosphere, and in some cases, a liquid sidestream. The impacts of all of these products must be considered in the design of the heat-drying facilities. Heat dried sludge should not be allowed to get wet, because moisture creates an environment favorable for regrowth of organisms. Once sludge gets wet again, anaerobic decomposition can begin with the concomitant generation of noxious odor. This is particularly a problem for sludges that have not been previously stabilized. Potential users of dried sludge prefer a granular or pelletized product. A product, which is dusty, odorous, or contaminated with materials such as plastics, strings, or cigarette butts, is difficult to sell or give away.

The gas stream exhausted from the dryers may be the source of pollution such as odor and visible emissions. The most effective control measure for this problem is afterburning. However, afterburning requires supplementary fuel and may be prohibitively expensive for many installations. Cyclones, wet scrubbers, electrostatic precipitators, and baghouses have been used with varying degrees of success. Wet-scrubbing, electrostatic precipitators and baghouses were tested for the control of odor and visible emissions from a Toroidal dryer located at the Blue Plains plant in Washington, DC. The electrostatic precipitator and wet scrubber were unable to reduce emissions sufficiently to satisfy Washington's stringent air pollution requirements. Baghouses were effective when operating, but they persistently caught fire as a result of ignited grease deposits and thus were not reliable.

Drying systems are exposed to heavy dusting and have had problems with fire. The combination of combustible particles, warm temperature, sufficient oxygen, and high-gas velocities make these systems susceptible to fire. Liquid sidestreams are produced by certain ancillary equipment in heat drying (e.g., wet scrubbers). These Sidestreams frequently can be recycled to the headworks of the treatment plant but may require separate treatment.

A major maintenance problem in some dryers is erosion of conveying equipment and drying of shells by the abrasive dried sludge. This is particularly a problem for dryers processing water activated sludge (WAS) from activated sludge plants, which has only coarse screening for grit removal. The use of ferric chloride as a dewatering aid may also create corrosive conditions that exacerbate the problem. Worn-out conveying equipment can lead to dusting problems. Abrasive sludge may result in replacement of rotary dryer drum shells every few years.

3.2.6. Energy and Costs

Thermal evaporation of water from sludge requires considerable energy. The amount of fuel required to dry sludge depends upon the amount of water evaporated. It is important that a dewatering step precede heat drying so that overall energy requirements can be minimized. Figure 2 shows a relationship between the solids content of the sludge and the energy required to produce a product containing 10% moisture. The energy estimates for heat drying of sludge must be considered in rough approximate, because values can vary considerably depending upon the type of dryer, whether or not energy recovery is a part of the process, the flow sheet, and the characteristics of the sludge.

The heat required to evaporate water from the wet sludge is made up of:

a. Heat to raise the sludge solids and associated residual water to the temperature of the sludge product as it leaves the dryer.
b. Heat to raise the water temperature to the point where it can evaporate and then to vaporize (latent heat).

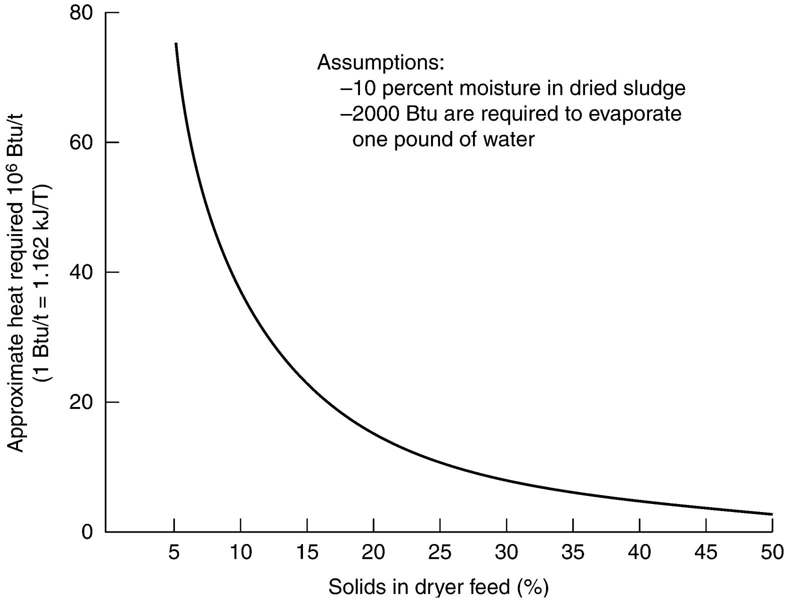


Fig. 2. Estimate of energy required to dry wastewater sludge as a function of dryer feed solids content. 1 t = 2000 lb; 1 T = 1000 kg. Source: US EPA.

c. Heat to raise the temperature of the exhaust gas, including water vapor, to the exhaust temperature.
d. Heat to offset heat losses.

The aforementioned heat must be supplied by the heating medium, for example, hot air or steam. Typical operating and maintenance costs for dewatering, drying, and bagging are shown in Table 5 (January 2002 cost).

4. DESIGN AND APPLICATION EXAMPLES

4.1. Example 1

About 10,000 lb/h (4540 kg/h) of a dewatered sludge containing 20% solids is to be dried by direct contact with hot air. The sludge temperature is 60°F (17°C). The temperature of the air prior to heating is 70°F (22°C), and its absolute humidity is 0.008 lb water/lb of dry air. The temperature of the dried sludge is 140°F (60°C). The dried sludge is 91% solids and 9% water. The dryer exhaust gas temperature is 240°F (116°C), and it contains 0.12 lb of water/lb of dry air. Radiant heat losses from the dryer structure are 1,000,000 Btu/h (1,054,000 kJ/h). A preheater is used to heat the air prior to its entering the dryer. The following heat capacity information is known or assumed:

Substance	Heat capacity (Btu/lb/°F)
Dry air	0.24
Dry solids	0.25
Water	1
Water vapor	0.45

Table 5
Estimated Costs for Sludge Dewatering and Drying[a]

Item	Annual cost (USD)	Cost/t (USD)[b]
Polymer	33,566	59.39
Gas	67,132	116.19
Labor	54,222	92.95
Power	28,402	51.64
Total	183,322	320.17

Source: US EPA (2).

[a] *Note:* January 2002 cost data (ENR CC Index = 6390.21).

[b] 1 t = 2000 lb.

Determine the following for a heat drying process system:

a. The required airflow (G).
b. The required air inlet temperature (t_2).
c. The evaporative efficiency.

Solution

Figure 3 is a schematic diagram for this example.

1. **Determination of the required airflow (G).** Calculate a moisture balance of substances entering and leaving the dryer.
   a. Moisture in:

$$\text{Moisture in sludge} = \left(10{,}000\frac{\text{lb sludge}}{\text{h}}\right)\left(0.8\frac{\text{lb water}}{\text{lb sludge}}\right) = 8000\,\text{lb/h (3.6 T/h)}$$

$$\text{Moisture in inlet air} = \left(G\frac{\text{lb dry air}}{\text{h}}\right)\left(0.008\frac{\text{lb water}}{\text{lb dry air}}\right) = 0.008\,G\ \text{lb/h}$$

   b. Moisture out:

$$\text{Moisture in sludge} = \left(10{,}000\frac{\text{lb sludge}}{\text{h}}\right)\left(0.2\frac{\text{lb dry solids}}{\text{lb sludge}}\right)\left(\frac{9\ \text{lb water}}{91\ \text{lb dry solids}}\right)$$
$$= 200\ \text{lb/h (91 kg/h)}$$

$$\text{Moisture in air} = \left(G\frac{\text{lb dry air}}{\text{h}}\right)\left(0.12\frac{\text{lb water}}{\text{lb dry air}}\right)$$
$$= 0.12G\ \text{lb/h}.$$

   c. Equate moisture in and moisture out

$$8000 + 0.008\,G = 200 + 0.12\,G.$$

   d. Solve for inlet airflow (G):

$$G = 69{,}600\ \text{lb/h (31.6 T/h)}.$$

   (Note: 1 T = 1000 kg.)

2. **Determination of the required air inlet temperature (t_2).** Calculate a heat balance for the dryer. A substance's heat content with respect to a given base temperature can be calculated, by assuming, the heat required to bring the substance from the base temperature

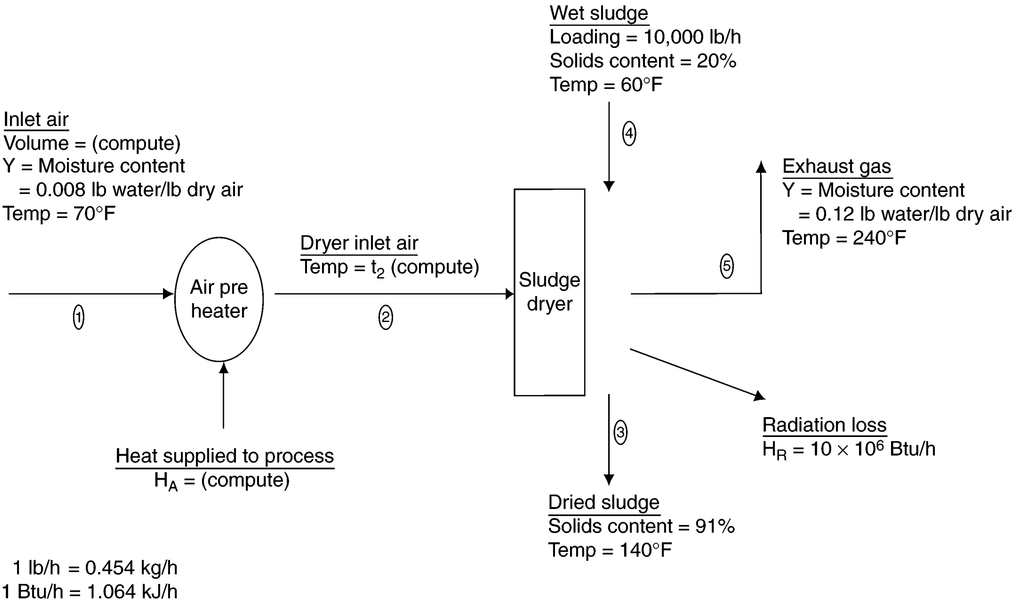


Fig. 3. Schematic for sludge drying example. Source: US EPA (2).

to the temperature being considered. For this example, a base temperature of 32°F (0°C) is arbitrarily selected, and heat content (also known as enthalpy) is calculated with respect to it. At steady-state, heat in must equal to heat out. Consider the heat content of streams entering and leaving the dryer:

a. Heat into the dryer is the sum of
   i. Heat content of sludge (H_4) entering the dryer (Fig. 3)
      - Heat content of dry solids

$$= \left(10{,}000\frac{\text{lb sludge}}{\text{h}}\right)\left(0.20\frac{\text{lb solids}}{\text{lb sludge}}\right)\left(0.25\frac{\text{Btu}}{\text{lb/°F}}\right)(60-32\text{°F})$$

$$= 14{,}000 \text{ Btu/h}.$$

      - Heat content of water

$$= \left(10{,}000\frac{\text{lb sludge}}{\text{h}}\right)\left(0.80\frac{\text{lb water}}{\text{lb sludge}}\right)\left(1\frac{\text{Btu}}{\text{lb/°F}}\right)(60-32\text{°F})$$

$$= 224{,}000 \text{ Btu/h}.$$

      - Summing

$$H_4 = 14{,}000 + 224{,}000 = 238{,}000 \text{ Btu/h}.$$

   ii. Heat content of air entering the dryer (H_2)
      - Heat content of dry air

$$= \left(69{,}600\frac{\text{lb}}{\text{h}}\right)\left(0.24\frac{\text{Btu}}{\text{lb/°F}}\right)(t_2 - 32\text{°F})$$

$$= 16{,}700\,(t_2 - 32) \text{ Btu/h}$$

- Determine the heat content of the moisture associated with the exhausted air. This includes heat required to raise the moisture temperature from 32°F (0°C) to the dewpoint, vaporize the moisture, and finally increase the vapor temperature to t_2. From psychrometric charts (6), the dewpoint (the temperature at which the air in question is saturated) of air containing 0.008 lb of water per pound of dry air is 50°F (10°C). From steam tables, the latent heat of vaporization at 50°F (10°C) is 1065 Btu/lb.

  Heat content of moisture associated with air

$$= \left(69{,}600 \frac{\text{lb dry air}}{\text{h}}\right)\left(0.008 \frac{\text{lb water}}{\text{lb dry air}}\right)$$

$$\left(\left[1 \frac{\text{Btu}}{\text{lb/°F}}\right](50-32\text{°F}) + 1065 + \left[0.45 \frac{\text{Btu}}{\text{lb/°F}}\right](t_2 - 50\text{°F})\right)$$

$$= 603{,}000 + 250.7\,(t_2 - 50)\ \text{Btu/h}$$

- Summing

$$H_2 = 16{,}700\,(t_2 - 32) + 603{,}000 + 250.7\,(t_2 - 50)$$

$$= 17{,}000\, t_2 + 55{,}600 \text{ Btu/h.}$$

b. Heat out of the dryer is the sum of

i. Heat content of the "dried" sludge (H_3)

- Heat content of dry solids

$$= \left(10{,}000 \frac{\text{lb sludge}}{\text{h}}\right)\left(0.20 \frac{\text{lb solids}}{\text{lb sludge}}\right)\left(0.25 \frac{\text{Btu}}{\text{lb/°F}}\right)(140 - 32\text{°F})$$

$$= 54{,}000 \text{ Btu/h.}$$

- Heat content of residual water

$$= \left(10{,}000 \frac{\text{lb sludge}}{\text{h}}\right)\left(0.20 \frac{\text{lb solids}}{\text{lb sludge}}\right)\left(\frac{9 \text{ lb water}}{91 \text{ lb sludge}}\right)\left(1 \frac{\text{Btu}}{\text{lb/°F}}\right) \times (140 - 32\text{°F})$$

$$= 21{,}400 \text{ Btu/h.}$$

- Summing

$$H_3 = 54{,}000 + 21{,}400$$

$$= 75{,}400 \text{ Btu/h.}$$

ii. Heat content of the exhausted air (H_5)

- Heat content of the dry air

$$= \left(69{,}600 \frac{\text{lb dry air}}{\text{h}}\right)\left(0.24 \frac{\text{Btu}}{\text{lb/°F}}\right)(240 - 32\text{°F})$$

$$= 3{,}474{,}000 \text{ Btu/h.}$$

- Determine the heat content of the moisture associated with the exhausted air. From psychrometric charts (6), the dewpoint of air containing 0.12 lb water per pound of dry air is 135°F (58°C). The latent heat of vaporization at 135°F (58°C) is 1017 Btu/lb.

Heat content of moisture associated with exhausted air

$$= \left(69{,}600 \frac{\text{lb dry air}}{\text{h}}\right)\left(0.12 \frac{\text{lb water}}{\text{lb dry air}}\right)$$

$$\left(\left[1 \frac{\text{Btu}}{\text{lb/°F}}\right]\left[135 - 32\text{°F}\right] + 1017 + \left[0.45 \frac{\text{Btu}}{\text{lb/°F}}\right]\left[240 - 135\text{°F}\right]\right)$$

$= 9{,}750{,}000$ Btu/h.

- Summing

$$H_5 = 3{,}474{,}000 + 9{,}750{,}000$$

$$= 13{,}224{,}000 \text{ Btu/h.}$$

iii. Radiant heat loss, $H_r =$ i.e., 1,000,000 Btu/h.

c. Calculate an overall heat balance around the dryer. At steady-state, heat into the dryer equal's heat out, $H_4 + H_2 = H_3 + H_5 + H_\text{r}$

Therefore: $238{,}000 + 17{,}000\ t_2 + 55{,}600 = 75{,}400 + 13{,}224{,}000 + 1{,}000{,}000$.

d. Solve for dryer inlet air temperature (t_2)

$t_2 = 826$°F (441°C).

3. **Determination of the evaporative efficiency.** Evaporative efficiency is equal to the of heat supplied to evaporate 1 lb of water, in comparison to the theoretical heat of vaporization:

a. Determine heat supplied to the process (H_A).

By an overall heat balance around the process (including the air preheater), $H_A = H_3 + H_5 + H_R - H_4 - \text{H}_1$.

i. From previous calculations,

"heat out" $= H_3 + H_5 + H_R$

$= 75{,}400 + 13{,}224{,}000 + 1{,}000{,}000$

$= 14{,}299{,}400$ Btu/h.

ii. From previous calculations,

$H_4 = 238{,}000$ Btu/h.

iii. Determine H_1, the heat content of the inlet air

- Heat content of dry air

$$= \left(69{,}600 \frac{\text{lb}}{\text{h}}\right)\left(0.24 \frac{\text{Btu}}{\text{lb/°F}}\right)(70 - 32\text{°F})$$

$= 635{,}000$ Btu/h.

- Heat content of moisture associated with dry inlet air

$$= \left(69{,}600 \frac{\text{lb dry air}}{\text{h}}\right)\left(0.008 \frac{\text{lb water}}{\text{lb dry air}}\right)\left[\begin{array}{l}\left(1 \frac{\text{Btu}}{\text{lb/°F}}\right)(50 - 32\text{°F}) + 1065 \\ + \left(0.45 \frac{\text{Btu}}{\text{lb/°F}}\right)(70 - 50\text{°F})\end{array}\right]$$

$= 608{,}000$ Btu/h.

- Summing

$$H_1 = 635{,}000 + 608{,}000 = 1{,}243{,}000 \text{ Btu/h}.$$

iv. $H_A = 14{,}290{,}000 - 238{,}000 - 1{,}243{,}000$

$= 12{,}809{,}000$ Btu/h.

b. Heat supplied to evaporate 1 lb of water.

$$= \frac{12{,}809{,}000 \text{ Btu/h}}{7800 \text{ lb water/h}} = 1642 \text{ Btu/lb of water.}$$

c. Heat by vaporization of water at the inlet sludge temperature = 1060 Btu/lb:

$$\text{Evaporative efficiency} = \frac{1060}{1642}(100) = 64\%$$

4.2. *Example 2*

Introduce a commercially available flash-drying equipment and its case history.

Solution

a. **Process equipment description.** Flash drying is the rapid removal of moisture by spraying or injecting the solids into a hot gas stream. The combustion engineering-Raymond flash drying and incineration process shown on Fig. 4, is typical of flash-drying units used in the United States.

The flash-drying process is based on three distinct components that can be combined in different arrangements. In the first component, the wet sludge cake is blended with previously dried sludge in a mixer to improve pneumatic conveyance. The blended sludge and the hot gases from the furnace at 1300°F (704°C) are mixed ahead of the cage mill, and flashing of the water vapor begins. Gas velocities on the order of 65–100 ft/s (20–30 m/s) are used. The cage mill mechanically agitates the sludge-gas mixture, and drying is virtually complete by the time the sludge leaves the cage mill. The mean residence time is a matter of seconds. The sludge, at this stage, has a moisture content of only 8–10% and is considered dry. The dried sludge is then separated from the spent drying gases in a cyclone. Temperature of the dried sludge is about 160°F (71°C), and the exhaust gas temperature is about 220–300°F (104–149°C). The dried sludge can be sent either to storage or to the furnace for incineration.

The second component is the incineration process. Gas, oil, coal, or partially dried sludge is burned in the furnace to provide heat needed to dry the sludge. Combustion air, provided by the combustion air fan, is preheated and injected into the furnace at high velocity to promote complete fuel combustion. Any ash that accumulates in the furnace bottom is periodically removed.

The third component is the effluent gas treatment facility or induced draft facility. This consists of the deodorizing preheater, the combustion air heater, the induced draft fan, and a gas scrubber. Odor is destroyed when the temperature of the gas from the cyclone is elevated in the deodorizing preheater. Part of the heat absorbed is recovered in the combustion air preheater. The gas then passes through a dust collector (generally a scrubber) and is discharged to the atmosphere.

Table 6 summarizes the US Environmental Protection Agency (US EPA) recommended monitoring variables, measurements, and instruments of the flash-drying process.

b. **Case study: Houston Texas.** The flash-drying operations at Houston, Texas, illustrate the operating experience and performance of the C-E Raymond Flash Drying process.

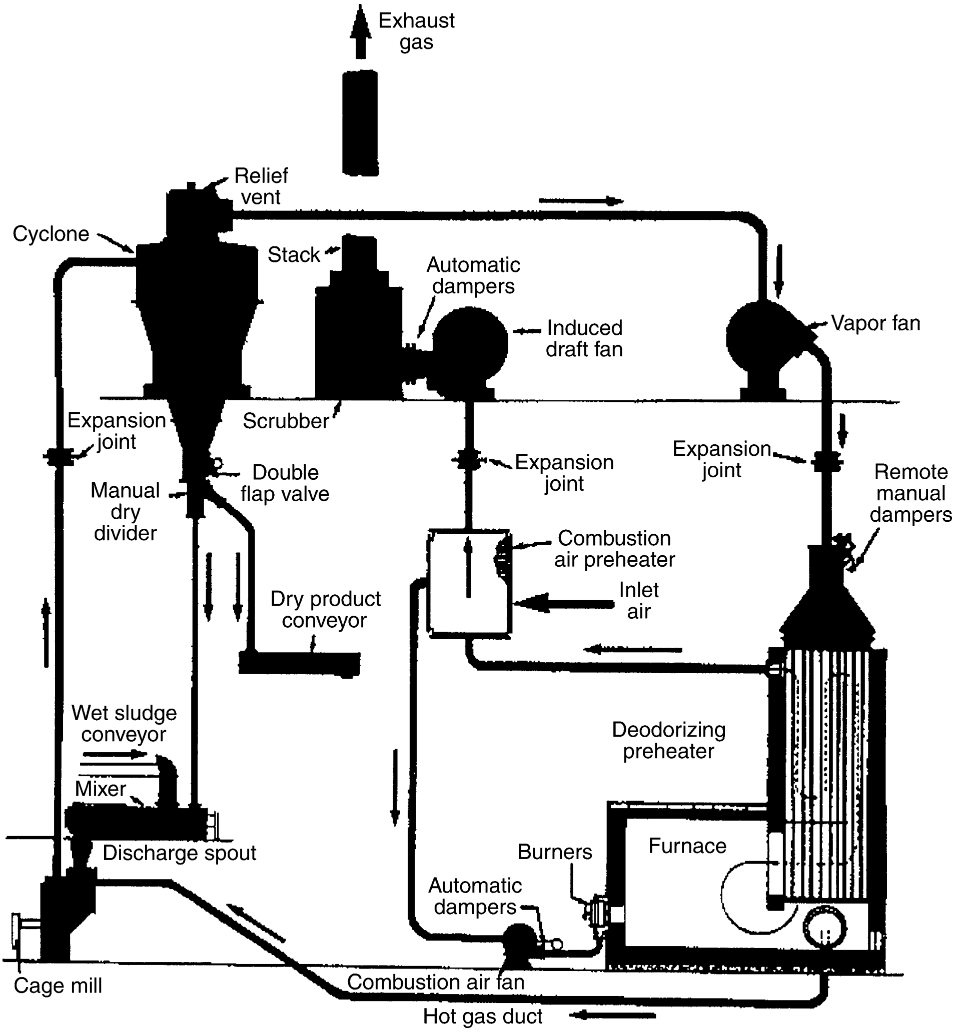


Fig. 4. Flash dryer system. Source: US EPA (2).

There are four flash dryers at the 45 MGD (1.97 m^3/s) Sims Bayou plant and seven flash dryers at the 75 MGD (3.29 m^3/s) Northside plant. The liquid process stream consists of bar screening and activated sludge. Sludge treatment consists of degritting, vacuum filtration with ferric chloride addition and flash drying.

After gravity thickening, the sludge solids concentration is about 2% at the Sims Bayou plant and about 3% at the Northside plant. The cake from the vacuum filters has about 15% solids. The ferric chloride additions amount to about 75 lb/t (37 kg/T) of dry solids, or about 3.8%.

Dewatered sludge is transported to the dryers by belt conveyors. Each flash dryer, with cage mill and 14 ft (4.3 m) diameter cyclone is rated at 12,000 lb of water/h (5450 kg/h) but is operated at 9000–10,000 lb of water/h (4090–4540 kg/h). Heat exchangers are provided for high temperature deodorization and for preheating the combustion air. The cage mill inlet temperature is 900–1150°F (480–620°C), and the temperature at the cyclone is about 220°F (104°C). The deodorization temperature is controlled around 1200°F

Table 6
Monitoring Variables, Measurements, and Instruments of Flash Drying Process

Process variable	Measurements	Instruments
Feed sludge	Flow, volume	Pump displacement Transport displacement
	Weight	Static Mass flow
	Moisture content	Portable ohmmeter Lab test
	Pipe empty	Capacitance Nuclear
Drying operation	Temperature	RTD (pad type)
Dried sludge	Flow, volume	Transport displacement
	Temperature	RTD (pad type)
	Weight	Static Mass flow
	Moisture content	Portable ohmmeter Lab test
Hot air furnace		
Burner operation	Flame monitoring	UV scanner
Fuel	Flow	Pitot tube Orifice Positive displacement
Combustion air	Flow	Pitot tube Orifice plate
	Pressure	Diaphragm Bellows
	Temperature	RTD
Heated Air	Temperature	Thermocouple
Fan monitoring	Flow loss	Vane Differential pressure Thermal
	Vibration	Accelerometer
Scrubber water		

Source: US EPA (3,18).
UV, ultra violet.

(650°C), and the stack gas temperature is 500–600°F (260–320°C) after heat recovery. The fuel used is natural gas, and the heat input is about 22 million Btu/h (23.2 million kJ/h) or 2200–2400 Btu/lb (5100–5600 kJ/kg) water evaporated.

Moisture content of the dried product is about 5.5%. About nine times as much solids on a dry weight basis are recycled to the predryer double paddle mixer, as are removed as product. The product is conveyed to a storage area or directly to railroad cars for shipment. The process is automated and panel boards are provided which indicate and record variables such as airflow, temperatures at critical points, and amperage on fan motors. The controls are enclosed in air-conditioned cubicles. Alarms indicate unsuitable temperature conditions. The controls for the ferric chloride feeding have proven to be inadequate and have led to operational problems.

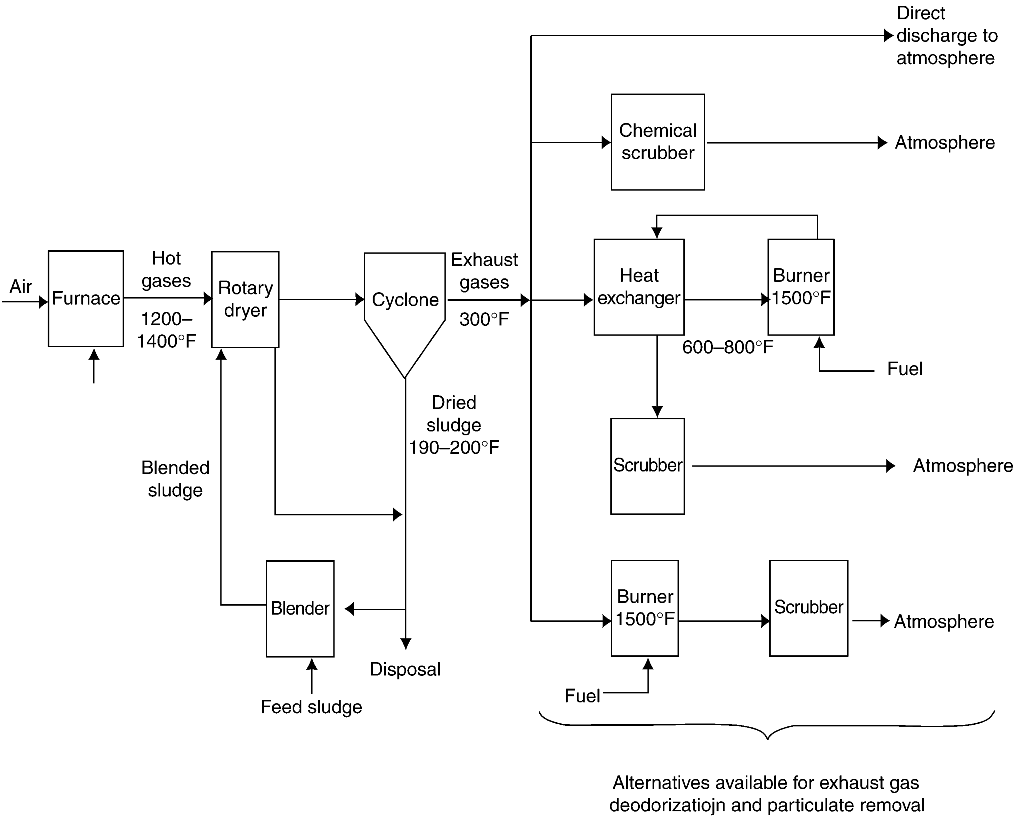


Fig. 5. Schematic for a rotary dryer. Source: US EPA (2).

Dust is also a major problem at the Sims Bayou plant. The dried sludge dust is extremely abrasive, causing wear on all mechanical equipment. Wet sludge has also overflowed the top of the conveyors at times, creating housekeeping problems. No specific cost data is available for the Houston facilities. The dried product, Hou-actinite, is sold through a broker by yearly contract.

4.3. Example 3

Introduce a commercially available rotary drying process equipment (known as direct rotary dryers) and its case history.

Solution

a. **Process equipment description.** Rotary dryers use a sloped rotating cylinder to move the material being dried from one end to the other by gravity. Direct, indirect, and direct-indirect rotary dryers have been used to dry sludge.

The features of a typical direct rotary drying system are illustrated on Fig. 5. Mechanically dewatered sludge is added to a mixer and blended with previously dried sludge to provide low moisture dryer-feed. Hot gas at temperatures of 1200°F (650°C) is added to the dryer, usually in a cocurrent flow pattern. After the sludge has been held in the dryer for 20–60 min, the dried sludge is discharged at 180–200°F (82–93°C). Exhaust gases are conveyed to a cyclone where entrained solids are separated from the gases. The spent gases exit at about 300°F (150°C). A portion of the dried product is recycled and the balance goes to a finishing step for further processing, or for disposal. Gaseous discharge from the cyclone goes to an air pollution control system for deodorization and particulate removal as necessary. Fig. 5 shows several alternatives for handling the exhaust gas. A long residence time in the dryer may minimize deodorization requirements.

The rotary drum usually consists of a cylindrical steel shell that revolves at 5–8 rpm. One end of the dryer is slightly higher than the other and the wet sludge is fed into the higher end. Flights projecting from the inside wall of the shell continually raise the material and shower it through the dryer gas, moving the material toward the outlet. Gas flow through the drum may be either cocurrent or countercurrent to the sludge flow. Gas velocities must be limited to 4–12 ft/s (1.2–3.7 m/s) to prevent dust from mixing with the exhaust gas. Table 7 summarizes the US EPA recommended monitoring variables, measurements, and instruments of the direct rotary drying process.

b. **Case study: Largo, Florida.** The Largo, Florida, Wastewater Treatment Plant has a rated capacity of 9 MGD (0.39 m^3/s) with average summer flow of 6 MGD (0.26 m^3/s) and winter flows greater than 9 MGD (0.39 m^3/s). The liquid process stream consists of coarse screening, grit removal, contact stabilization activated sludge, chlorination, and dual media filtration. Waste-activated sludge is aerobically digested, batch gravity decanted, and thickened. The thickened sludge is dewatered by belt filter presses and heat-dried in a rotary dryer. This system was supplied by Ecological Services Products, Inc. (ESP).

Approximately 1.6 dry t (1.45 T) of digested sludge is produced daily and is processed at a rate of 2.2 t (2 T)/d for a 5 d wk. Typical thickened aerobic sludge is 1–1.1% solids. The belt filter presses produce a sludge cake that is typically 10–12% solids. Polymer is used to condition the sludge prior to filtration.

The rotary dryer, manufactured by the Heil Company, has an evaporative capacity of approx 5400 lb water/h (2450 kg/h). The Heil dryer employs a 3-in-1-drum design. Sludge moves forward through the center cylinder, then back through the intermediate cylinder, and forward again through the outer cylinder toward a fan located at the discharge end of the machine. The three cylinders are concentric and are mechanically interlocked so that they rotate at the same speed. Internal–external flights on each cylinder repeatedly raise the sludge to the top of the drum. This design is claimed to provide better heat utilization by minimizing radiation losses, but maintenance on the drums is more complex than with a single shell.

The facilities were designed assuming 1000 lb/h (454 kg/h) of dry solids throughput, based on feeding sludge cake of about 20% solids. The dryer is water-limited because the cake produced by the belt has only 10–12% solids concentration. Actual throughput is about 600 lb/h (272 kg/h) of dry solids.

Heated air is provided by a natural gas burning furnace. Typical dryer inlet air temperature is about 800°F (427°C), and the outlet temperature is about 180°F (82°C). The average gas temperature in the dryer is estimated to be about 250°F (121°C). Off-gases from the cyclone separator are typically 120°F (49°C).

The dried product, Lar Grow, is a relatively fine pellet, produced naturally by the rotating drum. Product bulk density is 45–55 lb/ft^3 (720–880 kg/m^3). The bagged product moisture content is about 5%. The product is screened before bagging to remove cigarette filters and other nondegradable materials such as plastics. A garden products wholesaler has a contract to purchase the sludge produced at an average rate of 570 dry t/yr (517 T/yr). Because the wholesaler's markets are seasonal, the bagged product is stored on-site for a portion of the year. Typical operating and maintenance costs for dewatering, drying, and bagging are shown in Table 5 (adjusted to January 2002).

Table 7
Monitoring Variables, Measurements, and Instruments of Direct Rotary Drying Process[a]

Process variable	Measurements	Instruments
Feed sludge	Flow, volume	Transport displacement
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
		Lab test
Drying operation	Temperature	RTD (pad type)
	Speed	Reluctance
	Torque or power draw	Shear pin
		Ammeter
Dried sludge	Flow, volume	Transport displacement
	Temperature	RTD (pad type)
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
		Lab test
Hot air furnace		
Burner operation	Flame monitoring	UV scanner
Fuel	Flow	Pitot tube
		Orifice
		Vortex
		Positive displacement
Combustion air	Flow	Pitot tube
		Orifice
	Pressure	Diaphragm
		Bellows
	Temperature	RTD
Heated Air	Temperature	Thermocouple
Fan monitoring	Flow loss	Vane
		Differential pressure
		Thermal
	Vibration	Accelerometer
Scrubber water		

Source: US EPA (3,18).

Although a specific deodorization system has not been included, odor problems have been minimal. There is an occasional odor problem with when sludge that is too wet enters the dryer. There have been some problems with wear in the conveying facilities resulting of the dried sludge material being more abrasive than originally estimated. The pug mill blades and screw conveyor to the dryer have been replaced. Replacement parts have been specified to include heat treatment of the screw conveyor and the addition of cellite or carborundum plates on the wearing surfaces. The system supplier, ESP, has indicated that these changes will be considered for future equipment. There have been few other operating and maintenance problems.

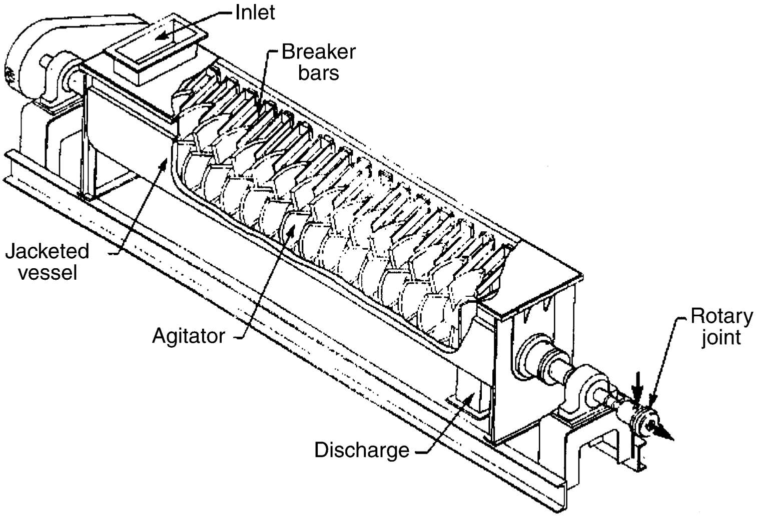


Fig. 6. Jacketed hollow-flight dryer.

4.4. Example 4

Introduce two types of commercially available rotary drying process equipment: (a) indirect rotary dryers and (b) direct–indirect rotary dryers. Also introduce a case history.

Solution

a. **Process equipment description of indirect rotary dryers.** Indirect dryers are not used in the United States for drying sludge. Vertical thin film dryers are used at the Dieppe, France coincineration facility (2,11). The two LUWA Double-Wall Dryers installed at Dieppe operate on 140 lb/in^2 (966 kN/m^2)-steam at a temperature of about 355°F (180°C). The evaporators are vertical, with top inlet and bottom outlet. Steam generated from refuse incineration is forced into the dryer and heats a "jacket" surrounding the incoming dewatered sludge. The sludge is spread over the inner cylindrical surface of the dryer by a rotor carrying self-adjusting vanes, at a top speed of about 25 ft/s (7.6 m/s). The water vapor travels upward, counter to the sludge flow, and is blown into the incinerator, where it is deodorized. The dried sludge falls onto a conveyor belt and is incinerated with the refuse.

Another type of indirect sludge dryer is the jacketed and/or hollow-flight dryer and conveyor. A schematic of a jacketed hollow-flight dryer is presented on Fig. 6. These units can perform the dual function of heat transfer and solids conveying in one piece of equipment—generally a horizontal, semicircular trough with a jacket or coil to provide heat (2,11). This equipment has one or more agitation devices (e.g., screw, flight, disc, and paddle) rotating on the axis through the center of the trough. A significant degree of agitation is necessary to maintain reasonable heat transfer. Simple screw conveyors fared notably poor in this regard, because increasing the speed reduces the residence time in the dryer by moving the sludge rapidly through the system. Heat transfer coefficients for this type of equipment range from 15 to 75 Btu/h/ft^2/°F (18.6–93 cal/h/cm^2/°C), depending on moisture content and degree of agitation.

The agitators, paddles, or flights should also be designed to minimize build-up on the walls of the dryer and on the agitator itself. Generally, baffles or ploughs should be provided between the flights to improve mixing and to break up any lumps that may

Table 8
Monitoring Variables, Measurements and Instruments of Indirect Rotary Drying Process, and Direct–Indirect Rotary Drying Process

Process variable	Measurements	Instruments
Feed sludge	Flow, volume	Transport displacement
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
		Lab test
Drying operation	Temperature	RTD (pad type)
	Speed	Reluctance
	Torque or power draw	Shear pin
		Ammeter
Dried sludge	Flow, volume	Transport displacement
	Temperature	RTD (pad type)
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
Hot air furnace		
Burner operation	Flame monitoring	UV scanner
Fuel	Flow	Pitot tube
		Orifice
		Vortex
		Positive displacement
Combustion air	Flow	Pitot tube
		Orifice
	Pressure	Bellows
		Diaphragm
	Temperature	RTD
Heated air	Temperature	Thermocouple
Fan monitoring	Flow loss	Vane
		Differential pressure
		Thermal
	Vibration	Accelerometer
Scrubber water		

Source: US EPA (3,18).

form. The rotating flights are often fitted with small paddles or similar projections to improve agitation and reduce fouling of the shell surface. Significant increases in heat transfer can also be obtained if the rotor is hollow and fitted for steam heating. A hollow heated rotor often provides one to two times the heat transfer area available in the shell.

b. **Process equipment description of direct–indirect rotary dryers.** The direct–indirect rotary dryer is similar to indirect dryers using hot air or gases as the heating medium. However, in direct–indirect drying, the heating medium is recirculated to flow in direct contact with the sludge in addition to heating the metal drying surfaces. Table 8 summarizes the US EPA recommended monitoring variables, measurements, and instruments for both of the above types, the indirect rotary dryers and the direct–indirect rotary dryers.

c. **Case Study.** The drying operation at Milwaukee's 200 MGD (8.76 m^3/s) Jones Island Plant employs 10 direct–indirect rotary and counterflow kiln-type dryers for treating waste-activated sludge. The plant is designed for continuous operation. To achieve this, nine dryers must always be in operation. The drying system produces over 74,000 t (67,300 T) of dried product. Thickened waste-activated sludge is conditioned with ferric chloride and filtered on vacuum filters. Wet filter cake (approx 14% solids) is mixed with an approximate equal weight of previously dried material in a screw conveyor and fed to the direct–indirect dryers. The 10 custom-built dryers are each 8 ft (2.4 m) in diameter and 57 ft (17.4 m) long. Each dryer can evaporate approx 10,000 lb (4540 kg) water/h (at 90% capacity) with an inlet air temperature of 1200°F (650°C). The rotating drum, with lifting angles, picks up the wet mixture that is dropped subsequently to the bottom as and when particles are shown. The sludge is continuously lifted and dropped through the hot gases, progressing as a moving curtain through the length of the dryer during the 45 min drying cycle. Since 1925 the granular dried sludge (Milorganite) has been sold as a fertilizer. Rejected dust and fine particles are pelletized, and the pellets are reground to produce granular saleable material.

The dryer air inlet temperature is controlled at 1200°F (650°C). The exhausted gas leaves the dryer at 250°F (120°C) and is passed through cyclone separators to remove fine particles. Each dryer has a separate furnace. Originally, coal was used as a fuel, then coke oven gas (after furnace modification), and then natural gas with standby fuel oil. In the mid-1970s, gas turbines were installed, and the gas from these turbines, at a temperature of approx 900°F (480°C), is now fed to the modified furnaces and two waste heat recovery boilers. The gas burners are used to provide the additional heat necessary to maintain the dryer inlet temperature at 1200°F (650°C). According to the records turbine exhausts heat supplies, 70% of the heat required for the sludge drying operation. The dried sludge product is abrasive, and the wet sludge is corrosive because of the ferric chloride used. Internals of the drum must be replaced about every 3 yr. Gobel (24) and Spraul (25) introduce more new dryer developments.

4.5. Example 5

Introduce the commercial process equipment (known as the Toroidal dryer) and its case history.

Solution

a. **Process equipment description.** The Toroidal (doughnut-shaped) dryer is a relatively new dryer that is used in the UOP, Inc. ORGANO-SYSTEM^R for sludge processing. The dryer works on a jet mill principle and contains no moving parts. Transport of solid material within the drying zone is accomplished entirely by high-velocity air movement. A simplified process flow diagram of the UOP ORGANO-SYSTEM^R is shown on Fig. 7. The system is made up of wet sludge storage, mechanical dewatering, sludge drying, air pollution control, final product finishing, and storage.

The mechanical dewatering step is designed to deliver the dewatered sludge to the dryer at about 35–40% solids. The dewatered sludge is mixed with previously dried sludge to reduce the moisture concentration of the dryer feed. Heated process air is distributed through three manifold jets to the lower segment of the toroidal drying zone chamber. The air from one of the three jets is directed in such a way as to impinge upon the incoming wet feed material and propel this material into the drying zone, where particle size reduction and drying begins. Additional jets in the drying zone convey the material into the toroid for additional drying, grinding, and classifying.

Process air and solids within the toroid move at a velocity of approx 100 ft/s (30 m/s). The high-velocity gas stream reduces the size of lumps or agglomerated feed material

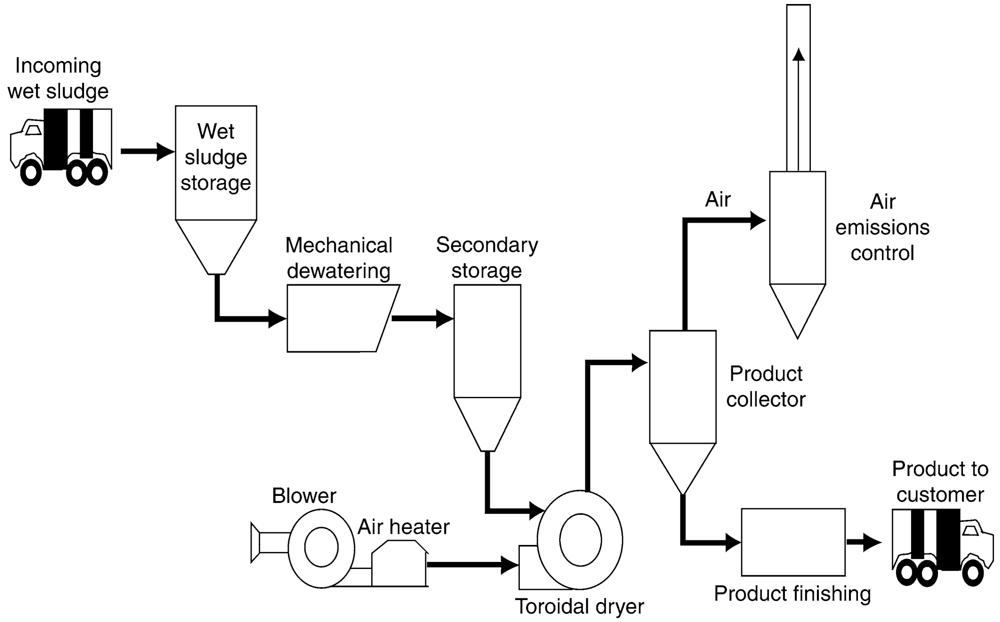


Fig. 7. Toroidal drying system. Source: US EPA (2).

by impingement against the interior walls of the drying chamber and by collision with other particles. Wetter and heavier particles travel a path along the internal periphery of the dryer, whereas drier and lighter particles are swept out with the gas stream and are removed from the drying zone. Heavy, wet particles stay in the dryer until they are broken up and dried.

The inlet temperature is usually controlled within the range of 500–1400°F (260–760°C). There is a sharp drop in the gas temperature within the dryer when the hot inlet gas stream meets the incoming wet sludge. The dryer exhaust temperature is usually controlled at a specific set point within the range of 190–300°F (90–150°C). The product temperature normally does not exceed 150°F (66°C). The dried sludge particles exiting the toroid are sent to a cyclone where they are separated from the gas stream. A portion of the dried sludge is again mixed with the wet feed, and the remainder is transferred to the product finishing section. There, the dried product may be extruded at a temperature of 140°F (60°C), cut into pellets, and bagged, if desired. Otherwise, the product is routed to subsequent sludge processes including codisposal/energy recovery or land application. Gases from the cyclone are treated by processes that may include scrubbers, electrostatic precipitators, and baghouses. Deodorizing chemicals may be required.

Table 9 summarizes the US EPA recommended monitoring variables, measurements, and instruments of Toroidal drying process.

b. **Case study: Washington, DC and West Chester, PL.** The toroidal dryer has been demonstrated on a full-scale basis. UOP Organic Recycling operated 240 t of water/d evaporative capacity ORGANO-SYSTEM^R at the Blue Plains wastewater treatment plant in Washington, DC, for over 3 yr. Raw sludge, digested primary sludge, and waste-activated sludge, as well as mixtures of these sludges, were processed. This system is no longer in operation. About 24 t of water/d vaporative capacity unit is installed at UOP's West Chester, Pennsylvania, research and development facility.

Table 9
Monitoring Variables, Measurements, and Instruments of Toroidal Drying Process

Process variable	Measurements	Instruments
Liquid of dewatered solids storage		
Feed sludge	Flow, volume	Transport displacement
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
		Lab test
Drying operation	Temperature	RTD (pad type)
Dried sludge	Flow, volume	Transport displacement
	Temperature	RTD (pad type)
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
		Lab test
Hot air furnace		
Burner operation	Flame monitoring	UV scanner
Fuel	Flow	Pitot tube
		Orifice
		Vortex
		Positive displacement
Combustion air	Flow	Pitot tube
	Orifice	
	Pressure	Bellows
		Diaphragm
	Temperature	RTD
Heated Air	Temperature	Thermocouple
Fan monitoring	Vibration	Accelerometer
	Flow loss	Vane
		Differential pressure
		Thermal
Scrubber water		

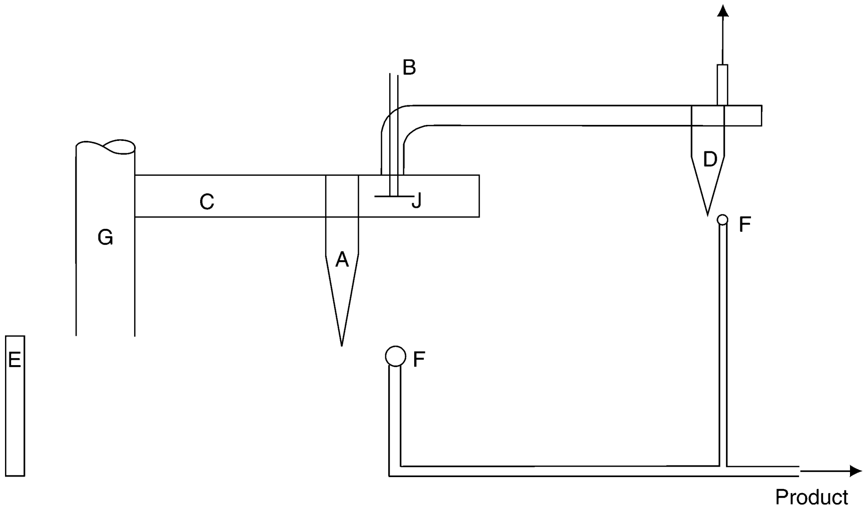
Source: US EPA (3,18).

4.6. Example 6

Introduce a prefabricated spray dryer, which is commercially available.

Solution

A spray-drying unit is shown schematically in Fig. 8. The dryer (A) consists of a cylindrical drying chamber tapering to a conical bottom for collection of the dried material. Slurry is fed into the dryer at (B) with the aid of a spray disk or nozzle. For slurries with significant amount of solids a spray disk rotating at 5000–10,000 rpm is recommended. The incoming slurry is atomized into tiny drops that move radially into a stream of hot gases, entering at the top of the dryer (C) and following a spiral motion downwards. The dried product is removed from the bottom of the conical chamber. A fan draws gases into a cyclone (D) to remove entrained solid particles (1).



A, dryer; B, slurry feed; C, hot gas entrance; D, cyclone;
E, furnace; F, rotary valve; G, stack; J, spray disc or nozzle

Fig. 8. A schematic diagram of a spray dryer unit (1).

The sprayed slurry from the nozzle or the spray disk, and the wet solid particles should not be allowed to strike the walls of the dryer before drying is complete. Therefore, large diameters, 8–30 ft, are common for the drying chamber. The residence time of the droplets in the drying chamber is only a few seconds (2–30 s), and the dried solids are not heated much above the wet-bulb temperature. Because of the low drying temperature and the short drying time, hot gases can be used in a spray dryer to dry beat-sensitive materials, such as food processing slurries and pharmaceuticals. Spray drying gives quite uniform, hollow particles, because surface drying outwards works faster than moisture diffusion in the interior of the particles (1).

The performance of a spray dryer depends on the residence time of the droplets in the drying chamber. The residence time is a complex function of the size and velocity of the droplets, the velocity and flow pattern of the drying gases, and the size and geometry of the drying chamber. Large drops may be under dried, while small ones may be over dried.

Spray drying has been applied to sludge treatment for drying, evaporation, pyrolysis, and incineration. Sludge is thickened to about 8% solids, ground to reduce the particle size, and atomized in the spray dryer, if the particles remain long enough in the dryer, above 600°F, ignition and incineration takes place. The Ansonia, Connecticut wastewater treatment plant processes dewatered sludge through a spray dryer with hot flue-gases (1300°F or 705°C) as the drying medium. The dried sludge with more than 90% solids is given away as a soil conditioner instead of being incinerated at the plant incinerator (1).

The city of Milwaukee investigated municipal sludge spray drying to obtain vitamin B_{12}. After washing the sludge, the extracted liquor was concentrated by vacuum evaporation before it was spray dried to obtain the vitamin. Atomized spray drying is also suitable for the treatment of food industry sludges to recover vitamins and proteins. Spray drying is a standard technique for recovering spent catalysts in the petrochemical industry.

Table 10
Monitoring Variables, Measurements, and Instruments of Spray Drying Process

Process variable	Measurements	Instruments
Feed sludge	Flow	Pump displacement
	Pressure	Bourdon or cylindrical seal
		Mass flow
	Density	Nuclear
		Optical
		Ultrasonic
Drying operation	Temperature	RTD (pad type)
Dried sludge	Flow, volume	Transport displacement
	Temperature	RTD (pad type)
	Weight	Static
		Mass flow
	Moisture content	Portable ohmmeter
Hot air supply	Temperature	Thermocouple

Source: US EPA (3,18).

Atomized spray drying provides flexibility of operation to either dry or incinerate sludge. However, equipment is expensive, and it is justified only if large amount of material are dried or when heat sensitive materials are processed. Sludge abrasiveness can cause material problems (wear and corrosion) with spray-drying equipment. Table 10 indicates the monitoring variables, measurements, and instruments recommended by the US EPA for monitoring a spray dryer.

4.7. Example 7

Introduce the manufacturers of commercial prefabricated dryers.

Solution

The manufacturers of commercial prefabricated dryers can be found in the following literature (14–17).

4.8. Example 8

Solution

a. The conventional heat conditioning process (Fig. 1) involves heating sludge (i.e., biosolids) to 140–240°C for short period of time under pressure of 250–400 lb/in^{2}g. It is essentially a thermal, nonchemical sludge conditioning process which prepares sludge for subsequent dewatering processes (such as vacuum filtration and filter presses without the use of chemicals). The amount of water removed from the sludge is negligible because the sludge is processed in a pressure reactor. It is not a drying process. However, the characteristics of sludge (i.e., biosolids) will be changed after heat conditioning.
b. In the heat drying process (Fig. 9) the moisture in the sludge is reduced by thermal treatment to 8–10% by the application of hot air, without combusting the solids materials. The sludge temperature is raised to 60–93°C, which is less than half the level required for sludge conditioning. For economic reasons, the moisture content of the sludge must be reduced as much as possible through mechanical means prior to heat drying. The five available drying techniques are flash, rotary, toroidal, and atomizing spray.

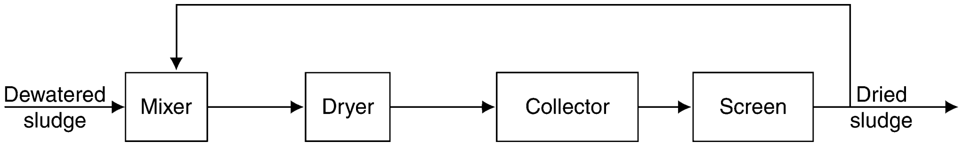


Fig. 9. Heat drying process. Source: US EPA (3).

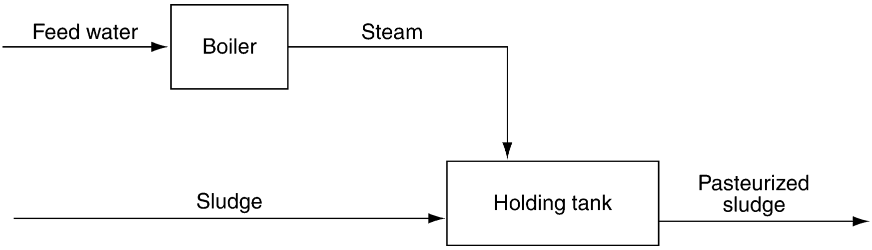


Fig. 10. Heat disinfection process (pasteurization process). Source: US EPA (3).

Figure 9 is a general flow diagram of a heat drying process, which is different from a heat conditioning process shown in Fig. 1. In a heat drying process, the sludge will not be processed under pressure, the water content will be reduced, and the sludge characteristics will not be changed significantly (18–22).

4.9. Example 9

Introduce (a) another low temperature thermal treatment process used mainly in Europe for sludge disinfection, and (b) an innovative heat conditioning process for disposal of low-level radioactive solid waste.

Solution

a. Heating to reach pasteurization temperatures is called “heat disinfection process” or “pasteurization process” and is a well known method of destroying pathogenic organisms, which has been applied successfully to disinfecting sludge. Pasteurization implies heating to a specific temperature for a time period sufficient to destroy undesirable organisms in sludge and make sludge more suitable for land disposal on cropland. Usually heat is applied at 70–75°C for a period of 20–60 min. Treatment can be applied to raw liquid sludge (thickened or unthickened) or digested sludge (3,23,26).
   Pasteurization, shown in Fig. 10, is usually a batch process, consisting of a reactor to hold sludge, a heat source, and heat exchanging equipment, pumping and piping and instrumentation for automated operation. Pasteurization has little effect on sludge composition or structure because sludge is only heated to a relatively moderate temperature.
b. An innovative heat conditioning process has been developed for disposal of low-level radioactive and infectious solid waste from hospitals, research institutions, universities, and so on (23). The innovative heat conditioning process involves thermal treatment under pressure, with digestion chemicals for converting radioactive solids waste to liquid sludge which can then be diluted with water for sewer discharge.

NOMENCLATURE

A	Area of wet surface exposed to drying medium (ft^2 [m^2])
A	Area of wet surface exposed to gas (ft^2 [m^2])

A	Surface area of heat transfer or sludge surface area exposed to radiant source (ft^2 [m^2])
h_c	Convective heat transfer coefficient (Btu/h/ft^2/°F [kJ/h/m^2/°C])
h_{cond}	Conductive heat transfer coefficient (Btu/h/°F [kJ/h/°C])
K_y	Mass transfer coefficient of the gas phase (lb water/h/ft^2/unit of humidity difference [kg/h/m^2/unit of humidity difference])
q_{conv}	Convective heat transfer (Btu/h [kJ/h])
q_{cond}	Conductive heat transfer (Btu/h [kJ/h])
q_{rad}	Radiation heat transfer (Btu/h [kJ/h])
t_g	Gas temperature (°F [°C])
t_m	Temperature of drying medium—for example, steam (°F [°C])
t_r	Absolute temperature of the radiant source (°R [K])
t_s	Temperature at sludge/gas interface (°F [°C])
t_s	Temperature of sludge at drying surface (°F [°C])
t_s	Absolute temperature of the sludge drying surface (°R [K])
W	Rate of drying (lb water/h [kg/h])
Y_a	Humidity of the gas phase (lb water/lb dry gas [kg/kg])
Y_s	Humidity at the sludge/gas interface temperature (lb water/lb dry gas [kg/kg])
ε_s	Emissivity of the drying surface (dimensionless)
σ	Stefan–Boltzman constant (1.73×10^{-9} Btu/h/ft^2/°R [4.88×10^{-8} kcal/m^2/h/K])

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10

Irradiation and Solid Substances Disinfection

Lawrence K. Wang, Paul Chen, and Robert Ziegler

CONTENTS

1. INTRODUCTION

1.1. Disinfection and Irradiation

Solid substances described in this chapter will include food, wastewater sludges and solid wastes. Disinfection is a process involving the destruction or inactivation of pathogenic organisms in the solid substances. The process is carried out principally to ensure sanitation or to minimize public health concerns. Destruction is the physical disruption or disintegration of a pathogenic organism, whereas inactivation, which is employed here, is the removal of a pathogen's ability to infect. An important but secondary concern may be to minimize the exposure of domestic animals to pathogens in solid substances. At the present time in the United States, the use of procedures to reduce the number of pathogenic organisms is a requirement before sale of sludges or recycled byproducts to the public as a soil amendment, or before recycling the sludges/byproducts directly to croplands, forests, or parks. Since the final use or disposal of sludges/byproducts may differ greatly with respect to public health concerns, and since a great number of treatment options effecting various degrees of pathogen reduction are available, the system chosen for reduction of pathogens should be tailored to the demands of a particular situation (1,2).

This chapter identifies the major pathogenic organisms found in wastewater sludges; briefly describes the pathogen characteristics, including size, life and reproductive requirements, occurrence in the solid substances, and survival under different environmental conditions; and discusses methods for reducing the number of pathogenic organisms in the sludge/byproducts. Various processes designed specifically for the reduction

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of pathogenic organisms in solid substances are only briefly discussed in this chapter because they are presented in other chapters of this handbook. However the high-energy irradiation process is introduced in this chapter in detail.

1.2. Pathogenic Organisms

A pathogen or pathogenic agent is any biological species that can cause disease in the host organism. The discussions in this chapter will be confined to pathogens that produce disease and complete their life cycles in humans in North America. These organisms or agents fall into four broad categories: viruses, bacteria, parasites, and fungi. Within the parasite category, there are protozoa, nematodes, and helminths. Viruses, bacteria, and parasites are primary pathogens that are present at some level in solid substances as a result of human activities. Fungi are secondary pathogens found in large numbers in solid substances when given the opportunity to grow during some treatment or storage process.

Pathogens enter environmental system from a number of sources:

a. Human waste, including feces, urine, and oral and nasal discharges.
b. Contaminated food and food waste from homes and commercial establishments.
c. Industrial waste from food processing, particularly meat packing plants.
d. Domestic pet feces and urine.
e. Biological laboratory waste such as those from hospitals.

In addition, where combined sewer systems are used, ground surface and street runoff materials, especially animal waste, may enter the sewers as storm flow. Vectors such as rats that inhabit some sewer systems may also add a substantial number of pathogens.

1.3. Pathogen Occurrence in the United States

Information on pathogen occurrence and associated morbidity and mortality data varies greatly with pathogenic species. Available data, complied by the Center for Disease Control (CDC) of the United States Public Health Service, indicates that enteric viral, bacterial, and parasitic infections annually affect tens of thousands of people in the United States (3–7). Data on the occurrence of bacterial disease in the United States is scarce. However, the frequent detection of enteropathic bacteria (bacteria which affect the intestinal tract), such as *Escherichia coli*, *Salmonella*, fecal *streptococci*, *Shigella*, and others in the contaminated food, untreated wastewater and wastewater sludges indicates that these pathogens and their associated diseases are endemic to the United States. Over 12% of stool samples checked by state and territorial public health laboratories were positive for one or more pathogenic parasites. *Ascaris lumbricoides*, which produces resistant ova, was found in more 2% of the samples (6,7). The frequent occurrence of enteric pathogens in the United States population indicates that pathogens should be expected in all wastewaters and sludges.

1.4. Potential Human Exposure to Pathogens

Man may be exposed to pathogens in wastewater sludge in a variety of ways and at greatly varying concentrations. Figure 1 lays out in simplified form some of the potential pathways. There is no firm scientific evidence to document a single confirmed case where human disease is directly linked to exposure to pathogens from wastewater

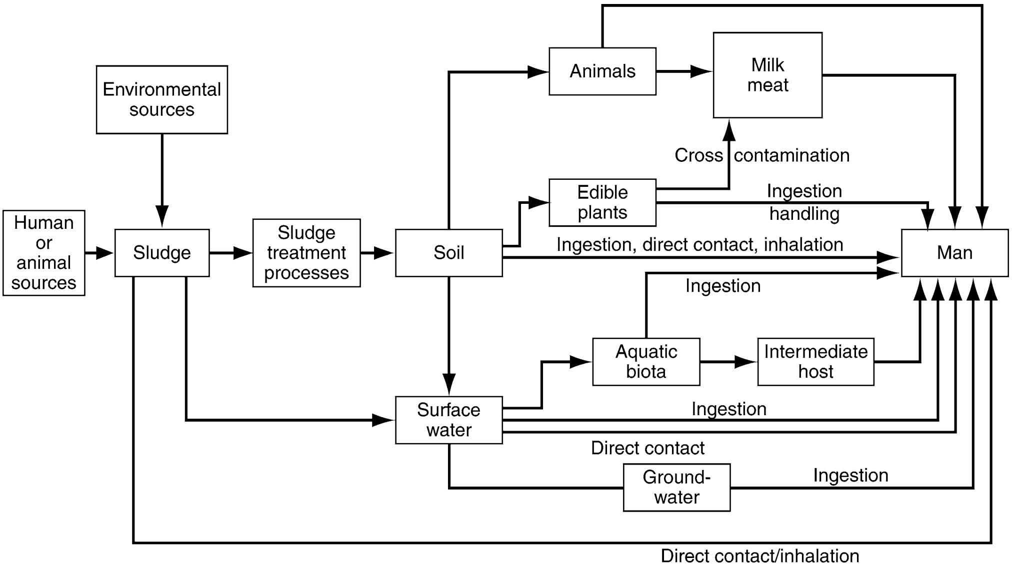

Fig. 1. Potential pathogen pathways to man.

sludge. However viable pathogens have been isolated from intermediate points in the sludge-management system, such as from surface runoff from sludge treated fields. These factors should be considered in the selection and design of a process for reducing the number of pathogenic organisms.

2. PATHOGENS AND THEIR CHARACTERISTICS

Viruses, bacteria, parasites, and fungi differ in size, physical composition, reproductive requirements, occurrence in the United States population, and prevalence in wastewater.

2.1. Viruses

Viruses are obligate parasites and can only reproduce by dominating the internal processes of host cells and using the host's resources to produce more viruses. Viruses are very small particles whose protein surface charge changes in magnitude and sign with pH. In the natural pH range of wastewater and sludges, most viruses have a negative surface charge. Thus, they will adsorb to a variety of material under appropriate chemical conditions. Different viruses show varying resistance to environmental factors such as heat and moisture. Enteric viruses are acid-resistant and many show tolerance to temperatures as high as 140°F (600°C). Many of the viruses that cause disease in man enter the sewers with feces or other discharges and have been identified, or are suspected of being, in sludge. The major virus subtypes transmitted in feces are as follows:

a. Adenoviruses.
b. Coxsackie virus, Group A.
c. Coxsackie virus, Group B.
d. ECHO virus (30 types).
e. Poliovirus (3 types).

f. Reoviruses.
g. Hepatitis virus A.
h. Norwalk agent.

Viruses are excreted by man in numbers several orders of magnitude lower than bacteria. Typical total virus concentrations in untreated wastewater are 1000–10,000 plaque-forming units (PFU)/100 mL; effluent concentrations are 10–300 PFU/100 mL. Wastewater treatment, particularly chemical coagulation or biological processes followed by sedimentation, concentrates viruses in sludge. Raw primary and waste-activated sludges contain 10,000–100,000 PFU/100 mL.

2.2. Bacteria

Bacteria are single-celled organisms that range in size from slightly less than 1 μ in diameter to 5 μ wide by 15 μ long. Among the primary pathogens, only bacteria are able to reproduce outside the host organism. They can grow and reproduce under a variety of environmental conditions. Low temperatures cause dormancy, often for long periods. High temperatures are more effective for inactivation, although some species form heat-resistant spores.

Pathogenic bacterial species are heterotrophic and generally grow best at a pH between 6.5 and 7.5. The ability of bacteria to reproduce outside a host is an important factor. Although solid substances may be disinfected, it can be reinoculated and recontaminated. Bacteria are numerous in the human digestive tract; man excretes up to 10^{13} coliform and 10^{16} other bacteria in his feces every day. The most important of the pathogenic bacteria are as follows:

a. *Arizona hinshawii.*
b. *Bacillus cereus.*
c. *Vibrio cholerae.*
d. *Clostridium pérfringens.*
e. *Clostridium tetani.*
f. *E. coli.*
g. *Leptospira* spp.
h. *Mycobacterium tuberculosis.*
i. *Salmonella paratyphi*, A, B, and C.
j. *S. seridai.*
k. *Salmonella* spp. *(>1500 types)*
l. *S. typhi.*
m. *Shigella* spp.
n. *Yersinia enterocolitica*
o. *Y. pseudotuberculosis*

2.3. Parasites

Parasite includes protozoa, nematodes, and helminths. Pathogenic protozoa are single-celled animals that range in size from 8 to 25 μm. Protozoa are transmitted by cysts, the nonactive and environmentally insensitive form of the organism. Their life cycles require that a cyst be ingested by man or another host. The cyst is transformed into an active organism in the intestines, where it matures and reproduces, releasing cysts in the feces. Pathogenic protozoa and the disease caused are listed in Table 1, together with the diseases they cause.

Table 1
Pathogen Occurrence in Liquid Wastewater Sludges

		Concentration (number/100 mL)	
Pathogen	Name or species	Unstabilized raw sludge[a]	Digested sludge[a,b]
Virus	Various	2.5×10^3 to 7×10^4	$100–10^3$
Bacteria	*Clostridia* spp.	6×10^6	2×10^7
	Fecal coliform	10^9	3×10^4 to 6×10^6
	Salmonella spp.	8×10^3	BDL[c]–62
	Streptococcus faecalis	3×10^7	4×10^4 to 2×10^6
	Total coliforms	5×10^9	6×10^4 to 7×10^7
	Mycobacterium tuberculosis	10^7	10^6
Parasites	*A. lumbricoides*	200–1000	0–1000
	Helminth eggs	200–700	30–70

[a] Type of sludge usually unspecified.
[b] Anaerobic digestion; temperature and detention times varied.
[c] BDL is less than detection limits (<3/100 mL).

Nematodes are roundworms and hookworms that may reach sizes up to 14 in. (36 cm) in the human intestines (1). The more common roundworms found in man are as follows:

1. Protozoa:
   a. *Acanthamoeba* spp.
   b. *Balantidium coli.*
   c. *Dientamoeba fragilis.*
   d. *Entamoeba histolytica.*
   e. *Giardia lamblia.*
   f. *Isosiora bella.*
   g. *Naegleria fowleri.*
   h. *Toxoplasma gordii.*
2. Nematodes:
   a. *Ancyclostoma dirodenale.*
   b. *Ancyclostoma* spp.
   c. *A. lumbricoides.*
   d. *Enterobius vermicularis.*
   e. *Necator americanus.*
   f. *Strongyloides stercoralis.*
   g. *Toxocara canis.*
   h. *T. cati.*
   i. *Trichusis trichiura.*
3. Helminths:
   a. *Diphyllobothrium latum.*
   b. *Echinococcus granulosis.*
   c. *E. multilocularis.*
   d. *Hymenolepis diminuta.*
   e. *Tymenolepis nana.*
   f. *Taenia saginata.*
   g. *T. solium.*

They may invade tissues other than the intestine. This situation is especially common when man ingests the ova of a roundworm common to another species such as the dog. The nematode does not stay in the intestine but migrates to other body tissue such as the eye and encysts. The cyst, similar to that formed by protozoa, causes inflammation and fibrosis in the host tissue. Pathogenic nematodes cannot spread directly from man to man. The ova discharged in feces must first embryonate at ambient temperature, usually in the soil, for at least 2 wk.

Helminths are flatworms, such as tapeworms, that may be more than 12 in. (30 cm) in length. The most common types in the United States (listed earlier) are associated with beef, pork, and rats. Transmission occurs when man ingests raw or inadequately cooked meat or the eggs of the tapeworm. In the less serious form, the tapeworm develops in the intestine, maturing and releasing eggs. In the more serious form, it localizes in the ear, eye, heart, or central nervous system.

2.4. Fungi

Fungi are single-celled nonphotosynthesizing plants that reproduce by developing spores, which form new colonies when released. Spores range in size from 10 to 100 μm. They are secondary pathogens in wastewater sludge, and large numbers have been found growing in compost (2). The pathogenic fungi, listed later, are most dangerous when the spores are inhaled by people whose systems are already stressed by a disease such as diabetes, or by immunosuppressive drugs. Fungi spores, especially those of *Aspergillus fumigatus*, are ubiquitous in the environment and have been found in pasture lands, hay stacks, manure piles, and the basements of most homes (2). The potential pathogenic fungi in wastewater sludge are as follows:

a. *Actinomyces* spp.
b. *Aspergillus* spp.
c. *Candida albicans.*

3. SOLID SUBSTANCES DISINFECTION

3.1. Long-Term Storage

Pathogen reduction has been recognized for years as a side benefit of sludge storage in lagoons. Hinesley and others have reported 99.9% reduction in fecal coliform density after 30 d storage (8). For an anaerobically digested sludge stored in anaerobic conditions for 24 wk at 39°F (4°C), Stern and Farrell (9) reported major reductions in fecal coliform, total coliform, and *Salmonella* bacteria. In similar tests at 68°F (20°C), the same bacteria could not be measured after 24 wk. Viruses were reduced by 67% at 39°F (4°C) and to less than the detectable limits at 68°F (20°C) in the same time period. Recent work by Storm and others showed fecal coliform reductions of one to three orders of magnitude during long-term storage of an anaerobically digested mixture of primary and waste-activated sludge (WAS) in facultative lagoons (10).

3.2. Chemical Disinfection

A number of chemicals used for wastewater sludge stabilization, including lime and chlorine, also reduce the number of pathogenic organisms in sludge.

3.2.1. Lime

Lime treatment of wastewater and sludge is a common practice (11). Plant-scale liming of wastewater sludge was evaluated at Lebanon, OH (12). Two chemical primary sludges, one with alum and one with ferric chloride, were limed to pH 11.5 and placed on drying beds. After 1 mo, *Salmonella* spp. and *Pseudomonas aeruginosa* were undetectable. Bench testing was also conducted on ferric chloride-treated wastewater raw sludges that were limed to pH 10.5, 11.5, and 12.5; these sludges were sampled after 0.5 and 24 h and bacterial tests were performed (12). Pathogenic bacteria reduction improved with time and was substantially better at pH values of 11.5 and 12.5. Qualitative checks for higher life forms such as *Ascaris* ova indicated that they survived 24 h at a pH greater than 11.0. Virus studies on limed sludges have not been reported, but a pH in excess of 11.5 should inactivate known viruses (9).

3.2.2. Chlorine

Chlorine is a strong oxidizing chemical used for disinfecting drinking water and wastewater effluents (13). It is effective for bacteria and virus inactivation if applied in sufficient quantity to develop a free chlorine residual in the solution being treated. Chlorine is less effective in disinfecting solutions with a high suspended solids concentration. Cysts and ova of parasites are very resistant to chlorine. Not much data is available on the potential of chlorine for reducing the number of pathogenic organisms in sludge. Some samples of sludge treated with large doses of chlorine in South Miami, Florida, and Hartland, Wisconsin, showed large reductions in bacteria and coliphages (14). Chlorine doses of 1000 mg/L applied to WAS with a 0.5% solids concentration reduced total bacteria counts by 4–7 logs and coliform bacteria and coliphage to less than detection limits. Primary sludge with a 0.5–0.85% solids concentration was treated with 1000 mg/L chlorine, and total and fecal coliform counts were reduced less than the detectable limits.

3.2.3. Quaternary Ammonium Compounds and Ozone

Other organic disinfectants, such as quaternary ammonium compounds (15–23), and strong oxidizing chemicals such as ozone (19–21,24–28) are sometimes used for solid substances disinfection. Today quaternary ammonium compounds are widely used in household kitchen, bathroom, and swimming pool disinfection. Oxygenation and ozonation have proven useful for sludge disinfection in both pilot plant and full scale municipal operations (22,24–26).

3.3. Low Temperature Thermal Processes for Disinfection

The number of pathogenic organisms in wastewater sludge can be effectively reduced by applying heat to untreated or digested sludges. Heat may be used solely for pathogen reduction as in pasteurization or as one step in a process designed to stabilize sludge, improve treatability or reduce mass. The focus of this section will be on sludge pasteurization. Other heating processes, such as thermal processing and incineration, are developed in other chapters of this handbook and will only be reviewed briefly here.

3.3.1. Sludge, Milk, and Foods Pasteurization

Man has recognized for many years that heat will inactivate microorganisms as well as the eggs and cysts of parasites. Different species and their subspecies show different sensitivities to elevated temperature and duration of exposure. Roediger, Stern, and Ward and Brandon (29–31) have determined the time and temperature relationships for disinfection of wet sludges with heat. Their results indicate that pasteurization at 158°F (70°C) for 30 min inactivates parasite ova and cysts and reduces population of measurable pathogenic viruses and bacteria less than the detectable levels. For bacteria, Ward and Brandon (31) found that fecal streptococci were most heat resistant, followed by coliforms and then *Salmonella*. Wang (23) indicates that a higher temperature for a shorter time period (195°F, 10 min) also destroys all pathogens.

3.3.2. Heat Conditioning

Heat conditioning includes processes where wet wastewater sludge is pressurized with or without oxygen and the temperature is raised to 350–400°F (177–240°C) and held for 15–40 min. These processes destroy all pathogens in sludge, and are discussed in another chapter in detail.

3.3.3. Heat Drying

Heat drying is generally done with a flash drier or a rotary kiln. Limited data from analyses on Milwaukee, Wisconsin's dried sludge Milorganite, produced with a direct-indirect rotary counter-flow kiln type dryer, indicates it is sterilized against bacteria (32). Data on samples of flash dried sludge taken in Houston, Chicago, Baltimore, and Galveston, showed no coliform bacteria in the Houston sludge and not more than 17 MPN/g dry sludge in the other sludges. Total nonconfirming lactose fermenters (spore formers) ranged from 14 to 240,000 MPN/g (33). No tests were made for viruses or parasites; other pathogens may also survive if some bacteria do.

Data for the Carver-Greenfield process gathered during testing by LA/OMA showed a seven order of magnitude reduction for total and fecal coliform, to a detectable level of less than one organism per gram (34). Fecal streptococci was reduced six orders of magnitude to two MPN/g and *Salmonella* from 50,000 MPN/g to less than 0.2 MPN/g. *Ascaris* ova were reduced to less than 0.2 ova/g.

3.4. High Temperature Thermal Processes for Disinfection

High temperature processes include incineration, pyrolysis, or a combination thereof (starved air combustion). These processes raise the sludge temperature more than 930°F (500°C) destroying the physical structure of all sludge pathogens and effectively sterilizing the sludge. The result of a high temperature process is sterilization unless short-circuiting occurs within the process. The readers are referred to another chapter of this handbook series for details.

3.5. Composting

Composting is considered here as a heat process because a major aim of sludge-composting operations is to produce a pathogen-free compost by achieving and holding a thermophilic temperature. Available data indicates that a well-run composting process greatly reduces the number of primary pathogens (14,35–40). However, windrow or aerated pile operations have not achieved a sufficiently uniform internal temperature to inactivate all pathogens. Adverse environmental conditions, particularly heavy rains, can significantly lower composting temperatures. Another chapter of this handbook series introduces the theory, principles, design considerations, operational considerations, and applications of composting process.

3.6. High Energy Radiation

The use of high energy radiation for solid substances disinfection has been considered for more than 25 yr. Two energy sources, β and γ-rays, offer the best potential system performance. β-rays are high energy electrons, generated with an accelerator for use in disinfection, while γ-rays are high energy photons emitted from atomic nuclei. As they penetrate, both types of rays induce secondary ionizations in the sludge. Secondary ionizations directly inactivate pathogens and produce oxidizing and reducing compounds that in turn attack pathogens. Sections 4 and 5 introduce the technologies in detail.

4. DISINFECTION WITH ELECTRON IRRADIATION

High energy electrons, projected through wastewater sludge by an appropriate generator, are being pilot tested as a means for inactivating or destroying pathogens in sludge at the Deer Island Wastewater Treatment Plant in Boston, MA (41). The electrons produce both biological and chemical effects as they scatter off material in the sludge. Direct ionization by the electrons may damage molecules of the pathogen, particularly the DNA in bacteria cell nuclei and the DNA or RNA of the viruses. The electrons also cause indirect action by producing hydrated electrons and H and OH free radicals that react with oxygen and other molecules to produce ozone and hydroperoxides. These compounds then attack organics in any solid substances, such as waste sludges—including pathogens—promoting oxidation, reduction, dissociation, and other forms of degradation.

The pathogen reducing power of the electron beam (e-beam) depends on the number and the energy of electrons impacting the sludge. E-beam dose rates are measured in rads; one rad is equal to the absorption of 4.3×10^{-6} Btu/lb (100 ergs/g) of material. As the radiation distributes energy throughout the volume of material regardless of the material penetrated, the degree of disinfection with an irradiation system is essentially independent of the sludge solids concentration within the maximum effective penetration depth of the radiation. The penetrating power of electrons is limited, with a maximum range of 0.2 in. (0.5 cm) in water or sludge slurries, when the electrons have been accelerated by a potential of one million volts (MeV).

For e-beam disinfection to be effective, some minimum dosage must be achieved for all sludge being treated. This effect is attained by using more than the average dosage desired for disinfection. One method used to ensure adequate disinfection is to limit the thickness of the sludge layer radiated so that ionization intensity of electrons exiting the treated sludge is about 50% of the maximum initial intensity. For the 0.85 MeV electrons used in the existing facility, this constraint limits sludge layer thickness to about 0.08 in. (0.2 cm). Accelerated electrons can induce radioactivity in substances which they impact. However, the electron energy levels for sludge irradiation, up to about 2 MeV, are well less than the 10 MeV needed to induce significant radioactivity with electrons.

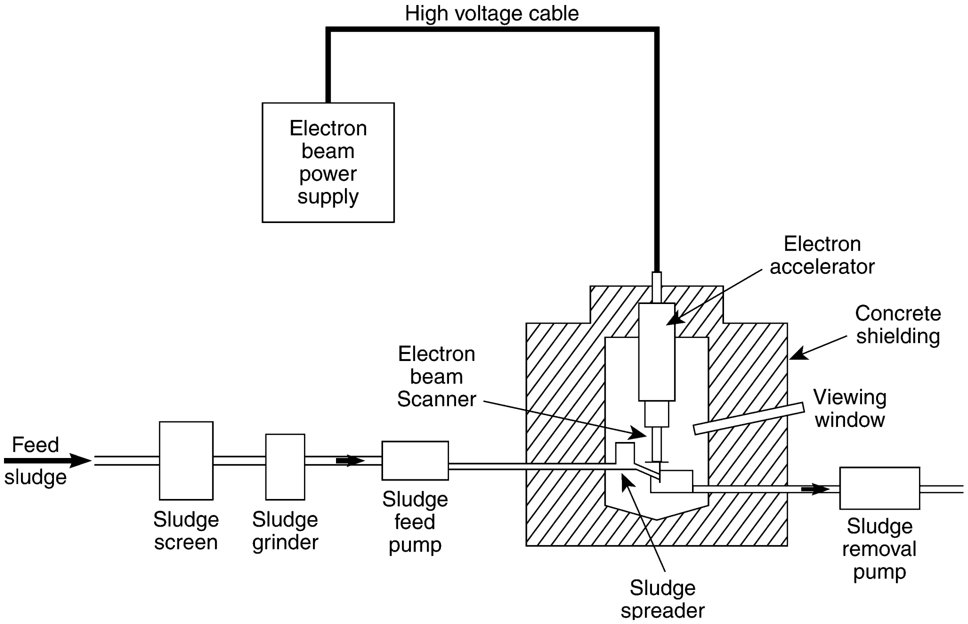

Fig. 2. Electron beam facility (*Source:* US EPA).

4.1. Electron Irradiation Process Description

Disinfection with an e-beam has been proposed for use on both untreated and digested sludges. The major system components of the Deer Island facility as shown in Fig. 2 include the sludge screener, sludge grinder, sludge feed pump, sludge spreader, electron beam power supply, electron accelerator, electron beam scanner, and sludge removal pump. A concrete vault houses the electron beam, providing shielding for the workers from stray irradiation, especially X-rays. X-rays are produced by the interaction of the electrons with the nucleus of atoms in the mechanical equipment and in the sludge. The pumps must be progressive cavity or similar types to assure smooth sludge feed. Screening and grinding of sludge prior to irradiation is necessary to assure that a uniform layer of sludge is passed under the e-beam. At Deer Island, sludge from the feed pump discharges into the constant head tank (*see* Fig. 3), which is equipped with an underflow discharge weir. Sludge is discharged under the weir in a thin stream and then flows down an inclined ramp. At the bottom of the ramp, it moves by free-fall into the receiving tank.

The electrons are first accelerated. They leave the accelerator in a continuous beam that is scanned back and forth at 400 times/s across the sludge as it falls free in a thin film from the end of the inclined ramp. The dosage is varied by adjusting the height of the underflow weir and hence the sludge flow rate. E-beam sludge irradiation must be considered a developing technology. The Deer Island irradiation facility is used for sludge disinfection. This project is designed to treat 0.1 MGD (4 L/s) sludge at up to 8% solids with a dosage of 400,000 rads. According to Shah, the facility has been operated about 700 h, since it was brought on line in 1976, with the longest continuous online time being 8 h (39).

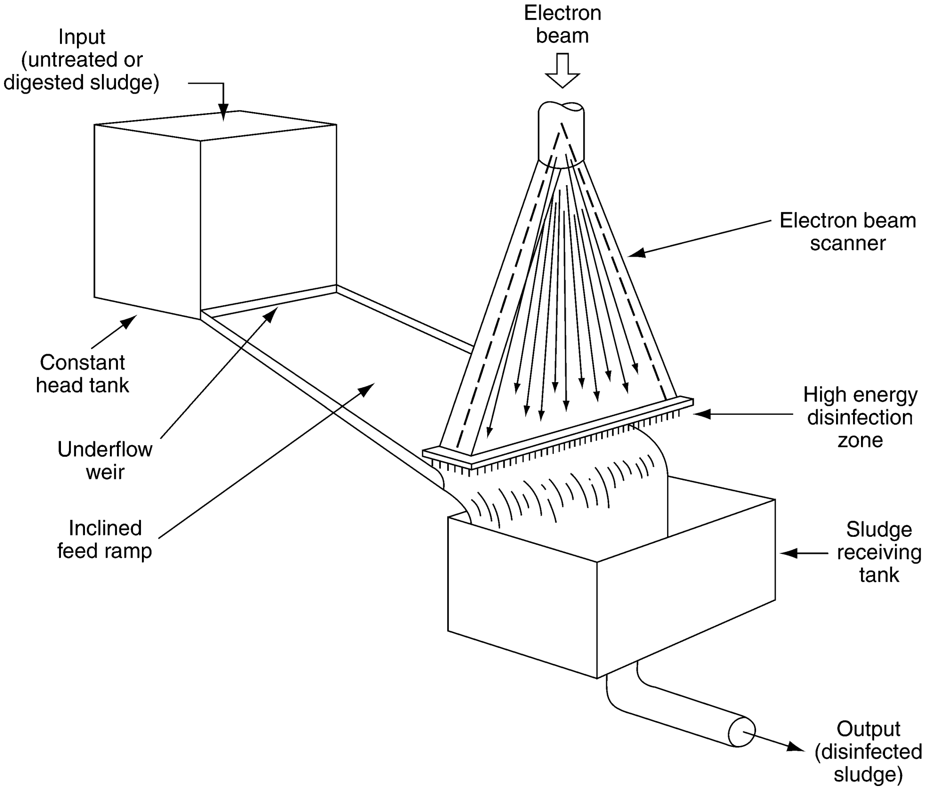

Fig. 3. Electron beam scanner and sludge spreader (*Source:* US EPA).

4.2. Electron Irradiation Design Considerations

Design criteria for an e-beam sludge facility is difficult to establish because operational data is available from only one pilot facility. However, the work at Deer Island provides good baseline information. A minimum level of electron irradiation should be 400,000 rads, which can best be supplied with a 1–2 MeV electron accelerator. This energy level provides good penetration for 0.2 in. (0.5 cm) thick sludge layers, making the achievement of a uniform sludge layer less important than with lower energy electrons. However, screening and grinding of sludge before disinfection are still necessary to ensure uniform spreading by this feed mechanism. The high energy electrons, combined with a short spacing of about 2.75 in. (7 cm) between the scanner window and the sludge film, ensure efficient energy transfer in the system. Only digested sludge has been irradiated at Deer Island. Nonstabilized sludge disinfection by e-beam irradiation still requires pilot-scale testing before any design is considered. Owing to the limited penetrating power of high energy electrons, this method of treatment is probably only feasible for liquid sludge disinfection. Piping pumps, valves, and flow meters should be specified as equal to those used for anaerobic sludge digestion systems.

4.3. Electron Irradiation Operational Considerations

Instrumentation needs for an e-beam facility should include flow measurement of, and temperature probes in the sludge streams entering and leaving the irradiator. Alarms as well as monitoring should be used to indicate variation in sludge flow and high or low radiation doses. Sludge disinfection by e-beam irradiation has large inherent flexibility. The radiation source (the e-beam) can be switched on and off as easily as an electric motor. The unit can be run as needed, up to its maximum throughput capacity. Electron accelerators have a proven record for reliability more than 20 yr in industrial applications and should prove dependable in wastewater treatment applications. The reliability of the electron beam generator and associated electronics presently used for medical and industrial applications is comparable with for the microwave radar systems at major airports (40). Accelerators for sludge disinfection would use the same basic components and would have similar reliability. Other system components, such as, pumps, screens, and grinders are all in common use in waste treatment plants. Cooling air for the scanner must be provided at several hundred cfm (about 10 m^3/s). This constant introduction of cooling air leads to the generation of ozone in the shielding vault around the accelerator. If the ozone were vented into the plant or into the atmosphere, some air pollution would result. At Deer Island, this problem is avoided by venting the cooling air through the sludge, where the ozone is consumed by chemical reduction. These reactions provide a small amount of additional disinfection and COD reduction.

Energy use for e-beam facilities has been estimated for the equipment used at Deer Island. A facility with a 50 kW (50 kJ/s) beam would require about 100 kW (100 kJ/s) of total electrical power including 25 W (25 kJ/s) for screening, grinding, and pumping, 10 kW (10 kJ/s) for window cooling, and 12 kW (12 kJ/s) for electrical conversion losses. Energy requirements for 0.1 MGD are 6 kWh/t (24 MJ/T) of wet sludge at 5% solids or 120 kWh/t (480 MJ/T) (39,41). Here 1 t = 2000 lb and 1 T = 1000 kg.

4.4. Electron Irradiation Performance

For untreated primary sludge, a dose of 400 kilorads (krads) with 3 MeV electrons reduced total bacteria count by 5 logs, total coliform by more than 6 logs, less than the detectable limits, and total *Salmonella* by over 4 logs, also below detectable limits. Fecal *streptococci* was only reduced by 2 logs with data indicating that some fecal streptococci are sensitive to radiation while others are resistant. For samples of anaerobically digested sludge irradiated at Deer Island with 0.85 MeV electrons, total bacteria was reduced by four logs at a dose of 280 krads, total coliform by 5–6 logs at a dose of 150–200 krads; a dose of 400 krads reduced fecal streptococci by 3.6 logs.

Virus inactivation has also been measured. A dose of 400 krads will apparently reduce the total virus measured as plaque forming units (PFU) by 1–2 logs. Laboratory batch irradiation of five enteric viruses showed about 2 logs reduction at a dose of 400 krads; Coxsackie virus was most resistant while Adeno virus was least resistant. These results correlate directly with virus size. Larger viruses are larger targets and hence more susceptible to electron "hits" (39,41).

Data for parasite reduction is scarce but 400 krads will apparently destroy all *Ascaris* ova (41). Comparing these performance data with information from Table 1 on the quantity of pathogens in sludge indicate that a dose of 400 krads may be adequate to disinfect anaerobically digested sludge, but raw sludge or aerobic sludge may require higher doses.

Odor problems are considerably lower for irradiated sludge as compared with pasteurized sludge (39). Irradiation of digested sludge with an e-beam may also improve sludge dewaterability and destroy some synthetic organic chemicals, as well as reduce pathogen levels. Irradiation has reduced specific resistance of sludge by up to 50% at a dose of 400 krads (39). Because specific resistance is normally measured on a log scale, a 50% reduction may indicate minimal improvement in sludge dewaterability. The energy requirements (fuel and electricity) for an irradiation system are estimated to be 90–98% less than those for heat pasteurization.

5. DISINFECTION WITH γ-IRRADIATION

5.1. γ-Irradiation Systems

5.1.1. γ-Irradiation System for Liquid Sludge

γ-Irradiation produces effects similar to those from an electron beam. However, γ-rays differ from electrons in two major ways. First γ-rays are highly penetrating. They require 25 in. (64 cm) thick water layer to stop 90% of the rays from a cobalt-60 (Co-60) source. In comparison of γ-ray 1 MeV electron can only penetrate about 0.4 in. (1 cm) of water. Second, γ-rays result from decay of a radioactive isotope. Decay from a source is continuous and uncontrolled; it cannot be turned off and on. The energy level (or levels) of the typical γ-ray from a given radioactive isotope is also relatively constant. Once an isotope is chosen for use as a source, the applied energy can only be varied with exposure time. Two isotopes, Cs-137 and Co-60, have been considered as "fuel" sources for sludge irradiators. Cs-137 has a half life of 30 yr and emits a 0.660 MeV γ-ray. It has been available in the United States as a byproduct from the processing of nuclear weapons waste. If the United States establishes a nuclear reactor spent-fuel rod reprocessing program, it would also be available at a rate of about 2 lb/t (1 kg/T) of fuel. Co-60 has a half life of 5 yr and emits two γ-rays with an average energy of 1.2 MeV. It is made by bombarding normal cobalt metal, which is stable cobalt isotope 59, with neutrons.

Two general types of γ-systems have been proposed for wastewater sludge disinfection (39). The first is a batch-type system for liquid sludge, where the sludge is circulated in a closed vessel surrounding the γ-ray source. Dosage is regulated by detention and source strength. The second system is for dried or composted sludge. A special hopper conveyor is used to carry the material for irradiation to the γ-ray source. Conveyor speed is used to control the dosage.

One γ-ray system in active operation is a liquid sludge facility at Geiselbullach (near Munich) in West Germany. Sludge has been treated there since 1973. The design capacity is 0.04 MGD (2 L/s) but the initial Co-60 charge only provided radiation to treat 0.008 MGD (0.3 L/s). The basic flow scheme is shown in Fig. 4. Digested sludge is pumped or otherwise moved into the vault with the Co source and circulated until the desired dosage is reached. The chamber is then completely emptied and recharged.

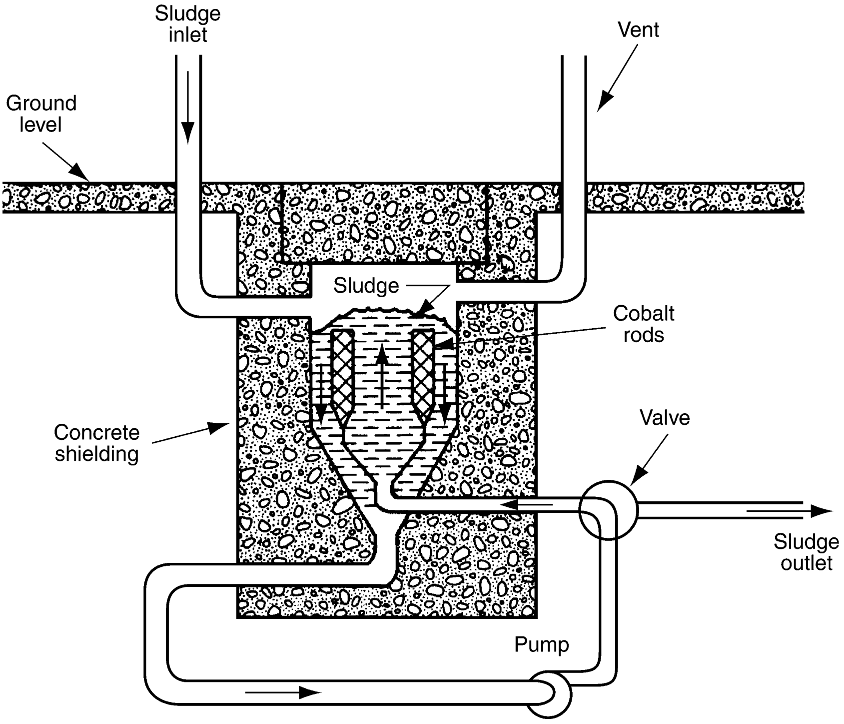

Fig. 4. γ-Irradiation facility (*Source:* US EPA).

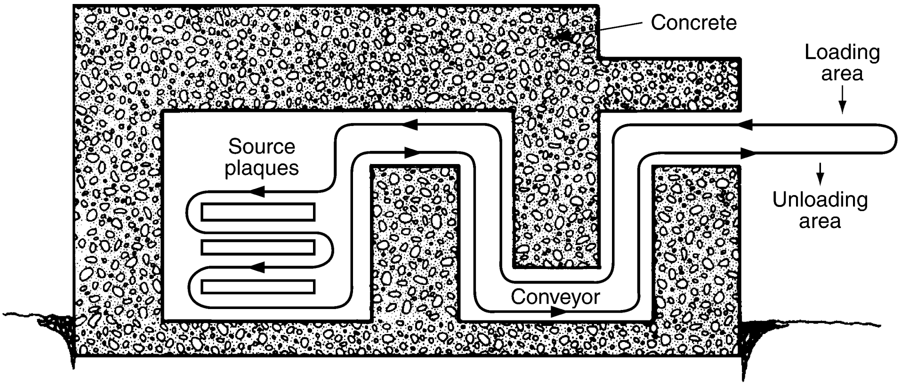

Fig. 5. γ-Irradiation facility for handling 25 t/d or more of dewatered sludge (*Source:* US EPA).

Wizigmann and Wuersching (42) reported on the efficiency of the Geiselbullach facility when the applied dose was 260 krads in 210 min. Bacterial tests were made on samples of processed sludge and showed a two logs reduction in total bacterial count, an Enterococcus reduction of two logs, and an *Enterobacteriaceae* reduction of 4–5 logs. Two of 40 samples were positive for *Salmonella*. Bacterial regrowth was measured in sludge-drying beds where the sludge was placed after irradiation.

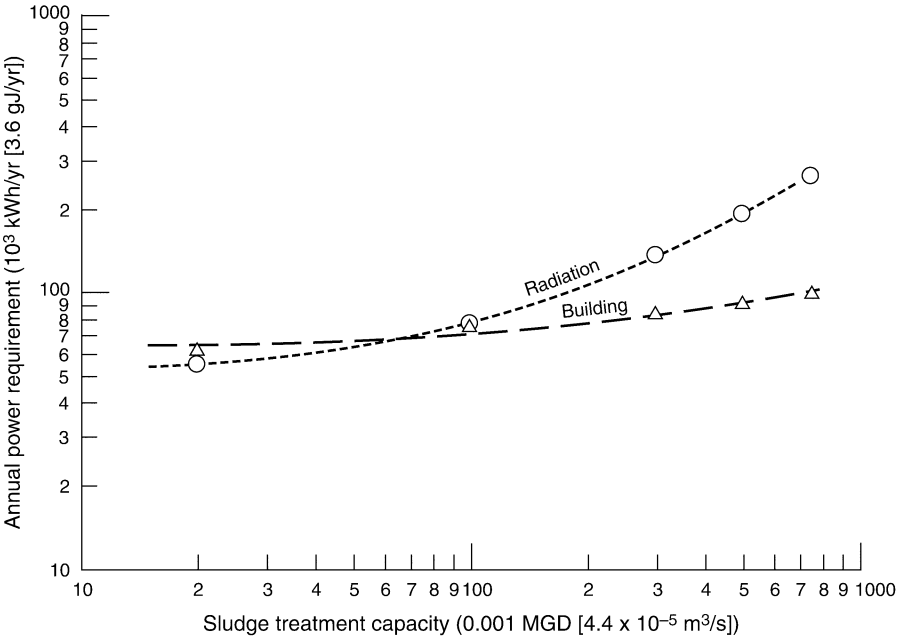

Fig. 6. Annual power requirement for γ-irradiation treatment of liquid sludge (*Source:* US EPA).

Plastic encapsulated bacteria samples were also irradiated in the system to a dosage of 260 krads. Two of nine *E. coli* strains were radiation-resistant and reduced 5–6 logs; three strains were totally inactivated, and four strains were reduced 6–8 logs. Tests on 10 strains of *Salmonella* in 170 samples showed 4–7 logs reduction, with 85% of the samples more than 5 logs and 61% more than 6 logs. *Kiebsiella* were reduced 6–8 logs. Gram-negative species were more sensitive to γ-radiation than Gram-positive ones, and spores were more resistant than vegetative forms. A comparison of the disinfection results of the real sludge samples and the plastic-encapsulated cultures indicates that circulation in the sludge system apparently did not result in a very uniform dose exposure.

Parasite ova (*Ascaris suum*) circulated through the system in plastic capsules failed to develop during 3 wk of incubation. This observation period was not adequate; however, to conclude that long-term recovery would not take place. Land spreading of the sludges treated at Geiselbullach has been well received by local farmers and the general public. No radiation hazards have resulted and the treated sludges satisfy disinfection requirements. The competing system in Germany, heat pasteurization, requires more energy and produces an odorous product that is more difficult to handle.

5.1.2. γ-Irradiation System for Dried or Composted Sludge

A dry sludge irradiation system using a γ-source has been developed by Sandia Laboratories in Albuquerque, New Mexico. The 8 t/d (7.2 T/d) facility contains about 1 million Ci of Cs-137. The facility is used to irradiate bagged composted sludge for agricultural experiments and bagged dried raw primary sludge for testing as a cattle-feed supplement. Owing to the high cost of Co-60, the overall viability of any sludge

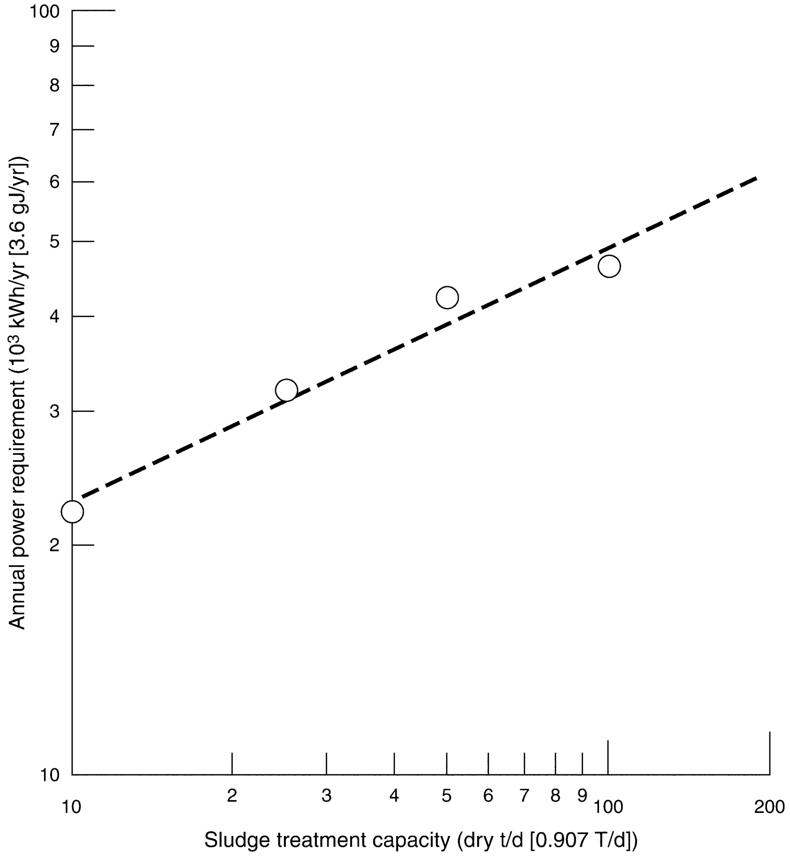

Fig. 7. Annual power requirement for γ-irradiation treatment of dewatered sludge (*Source:* US EPA).

irradiation facility in the United States depends on Cs-137 supplies. Cs-137 will be available in quantity only if the political and technical difficulties associated with power plant fuel rod reprocessing can be resolved. About 200 MCi of Cs-137 could be available from processing waste from weapons manufacture and could be used for further testing. The bucket conveyor used by the dry system to move the sludge past the radiation source is shown in Fig. 5.

5.2. γ-Irradiation Design Considerations

5.2.1. General Design Considerations

The design criteria for γ-irradiation facilities depend on the type of wastewater sludge treated. Current literature discussions suggest a dose of 400 krads but this level does not ensure complete virus removal (39). The dosage level should probably be varied in relation to other treatments the sludge receives. A composted, bagged product with 80% solid content needs a lower dose than a mixture of raw primary and WAS because the dried product already has a reduced pathogen level owing to the drying process. The storage capacity for both untreated and irradiated sludge should be equal to that of a pasteurization facility of similar size.

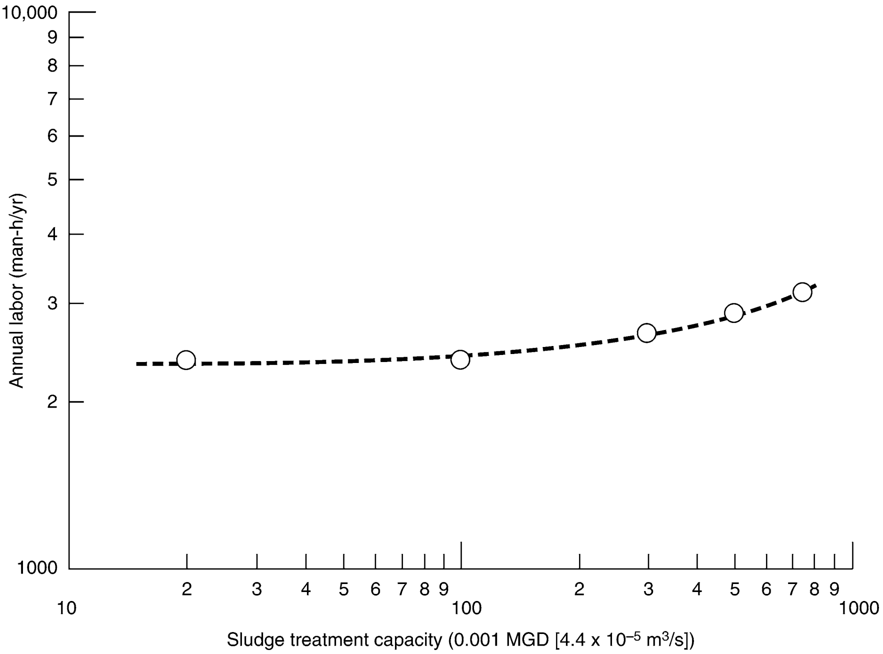

Fig. 8. Annual labor requirement for γ-irradiation treatment of liquid sludge (*Source:* US EPA).

When a dry system radiation source is not in use, it should be shielded in a steel-lined concrete vault. The vault should be designed to be flooded with water during loading and unloading of the radiation source, to shield workers from radiation. Provision must be made for pool water treatment in the event that the radiation source leaks. Cooling air is circulated around the source both during system operation and down times. This air must be filtered to prevent a radioactive air release. Because the dried sludge is a flammable material, there must be smoke and/or heat detection and a fire suppression system. For a liquid storage system the treatment vessel serves as a radiation source storage vault.

5.2.2. Facility Design Components

The liquid facility may include the following components:

a. Insulated concrete building with 25 ft (7.6 m) ceiling.
b. Equalization sludge storage tank.
c. Emergency water dump tank (for source shielding water).
d. Irradiating capsules (radiation source).
e. Steel-lined source handling pool.
f. Deionizer.
g. Data acquisition and control system.
h. Oxygen injection facility.
i. Pumps, piping, and flow meters.
j. Radiation alarm.
k. Fire suppression system.

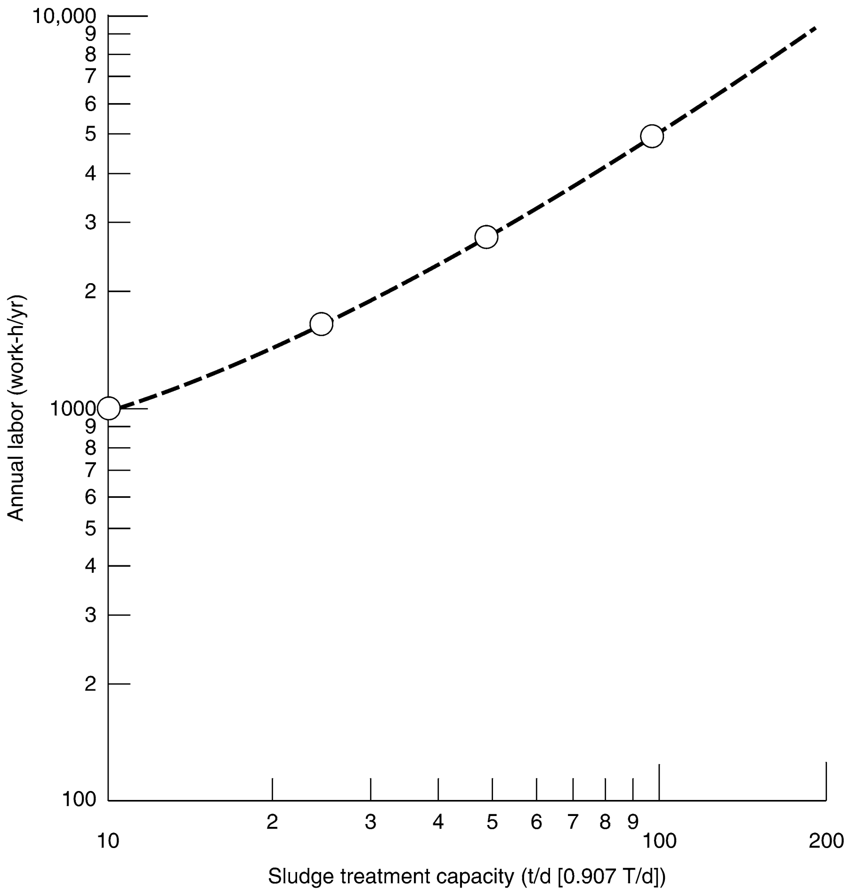

Fig. 9. Annual labor requirement for γ-irradiation treatment of dewatered sludge (*Source:* US EPA).

The dry system (Fig. 5) uses a bucket conveyor to move the sludge past the radiation source. This dry system may include the following:

a. Loading and unloading conveyors.
b. Concrete shielding.
c. Source handling pool.
d. Holder for the Cs-137 capsules.
e. Holder moving mechanism.
f. Steel building.
g. Pumps.
h. Ventilators.
i. Hoists.
j. Radiation alarm system.
k. Pool water testing tank.
l. Fire suppression system.

5.3. γ-Irradiation Operational Considerations

Instrumentation should include radiation detectors and flow metering for the wet sludge system. When either facility is operating, arrangements must be made for periodic radiation safety inspection. The disinfection effectiveness should also be tested by periodic sampling of the sludge before and after disinfection. Ahlstrom and McGuire (43) projected annual energy requirements for both wet and dry γ-irradiation facilities, using a dose rate of 1000 krads. Their results are summarized on Figs. 6 and 7. For a 0.1 MGD (4 L/s) facility treating sludge with 5% solids, 300 d/yr, the unit energy use is about 5.2 kWh/1000 gal (5 MJ/m^3) or 25 kWh/t (100 MJ/T) dry solids. For a plant treating 35 t/d (32 T/d) at 60% solids, 300 d/yr, the energy use is 5.6 kWh/t dry (22 MJ/T) solids, almost 80% less than the facility treating 5% solids. These energy uses should be compared with 120 kWh/t (450 MJ/T) for an e-beam system. The liquid system requires a much larger Cs-137 charge because it will be treating almost 12 times the volume of material at the same dose level. However, the rod configuration for a dry facility would be much less efficient in terms of radiation transfer than a liquid one. (43,44,50).

Figure 8 shows the annual labor requirements for γ-irradiation treatment of liquid sludge, while Fig. 9 shows the annual labor requirements for γ-irradiation treatment of dewatered sludge.

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11

Inorganic Chemical Conditioning and Stabilization

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

Conditioning involves the chemical and/or physical treatment of biosolids to enhance water removal and improve solids capture. The three most common conditioning systems use inorganic chemicals, organic polymers (covered in another chapter), or heat. Table 1 shows and compares the effects of conditioning processes on a biosolids mixture of primary and waste-activated sludge (WAS).

Conditioning always has an effect on the efficiency of the dewatering process that follows (1–5). Any evaluation of the conditioning process must therefore, take into consideration capital and operating and maintenance costs for the entire system. These costs include the impact of sidestreams on other plant processes, the plant effluent and resultant air quality. Some treatment plants are required to remove phosphorus, although less of these plants must do so today than many years ago. This is both because of the bans on the use of detergents containing phosphorus and the generally decreased use of phosphorus in household products. Phosphorus is often removed by the addition

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Table 1
Effects of Conditioning on a Biosolids Mixture of Primary and WAS

	Inorganic chemicals	Organic polymers	Heat
Conditioning mechanism	Coagulation and flocculation	Coagulation and flocculation	Alters surface properties and ruptures biomass cells, releases chemicals, hydrolysis
Effect on allowable solids loading rates	Will increase	Will increase	Will significantly increase
Effect on supernatant stream	Will improve suspended solids capture	Will improve suspended solids capture	Will cause significant increases in color, suspended solids, soluble BOD, NH_3–N, and COD
Effect on manpower	Little effect	Little effect	Requires skilled operators and a strong preventive maintenance program
Effect on sludge mass	Significantly increases	None	Reduces present mass but may increase mass through recycle

Source: US EPA.

of chemicals, including ferric chloride, aluminum sulfate (alum), sodium aluminate, and lime with some kind of polyelectrolyte to facilitate coagulation and settling (6). Because of the enormous materials handling difficulties, the interest in using lime has declined.

Precipitating phosphorus can as much as double the amount of biosolids requiring treatment and disposal. The amount of chemicals that must be added is a function of the amount of phosphorus needing removal. Fortunately, with the quantity of phosphorus in wastewater diminishing, this quantity is decreasing. If precipitating phosphorus removal is contemplated, laboratory/pilot tests should be performed to determine the mass and volumes of biosolids to be expected, the degree to which the biosolids can be thickened, and how well the biosolids can be conditioned and dewatered if appropriate. While in every instance there will be more biosolids to contend with (an additional 30–100%), the biosolids will thicken and dewater differently. Lime biosolids readily thicken and dewater, whereas hydroxide biosolids (ferric chloride, alum, sodium aluminate) usually thicken and dewater poorly requiring considerable conditioning with polymers. In fact the best analogy is that difficult-to-handle hydroxide biosolids behave like a poor quality activated sludge. It is also usual to find hydroxide biosolids that will at best dewater to a final cake solids concentration of 10–15% on a centrifuge, belt press, and vacuum filter (1–3). These "still wet" biosolids may be hard if not impossible to lift and unacceptable for disposal in landfill.

2. FACTORS AFFECTING BIOSOLIDS CONDITIONING

Wastewater solids are made up of screenings, grit, scum, and biosolids. Wastewater biosolids consist of primary, secondary, and/or chemical solids with various organic and

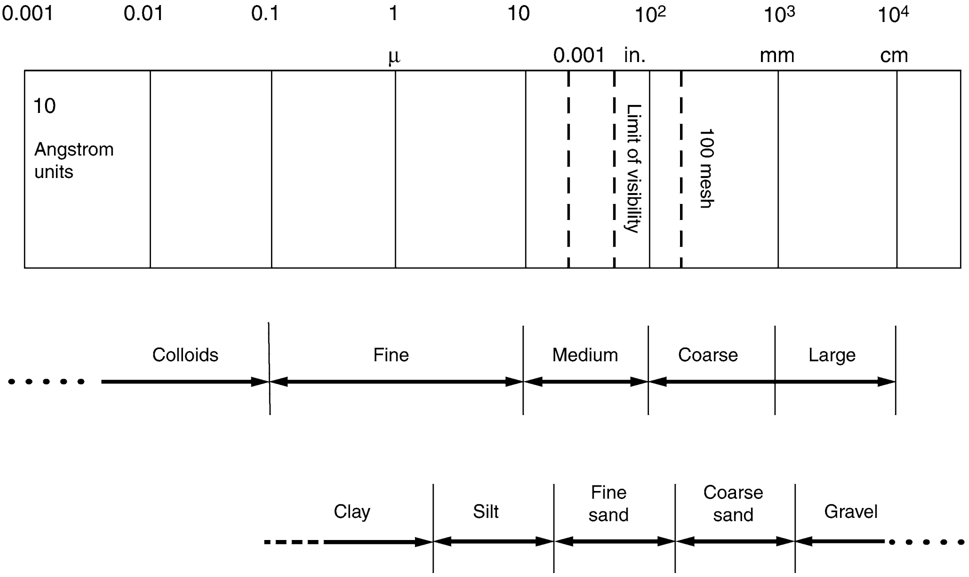

Fig. 1. Particle size distribution of common materials (*Source*: US EPA).

inorganic particles of mixed sizes. Biosolids have various internal water contents, degrees of hydration, and surface chemistry. Biosolids characteristics that affect dewatering (for which conditioning is used) are particle size and distribution, surface charge and degree of hydration, and particle interaction. Furthermore, such things as biopolymer production, degree of filamentous growth, primary to secondary biosolids ratio, and inorganic content also play an important role (5).

Particle size is considered to be the single most important factor influencing biosolids dewaterability. As the average particle size decreases, primarily from mixing or shear, the surface/volume ratio increases exponentially (4). Increased surface area means more hydration, higher chemical demand, and increased resistance to dewatering. Figure 1 shows relative particle sizes of common materials.

Raw municipal wastewater contains a significant quantity of colloids and fines that because of their size, 1–10 µm, will almost all escape capture in primary clarifiers if coagulation and flocculation are not used. Secondary biological processes, in addition to removing BOD, also partially remove these colloids and fines from wastewater. As a result, biological biosolids, especially WAS, are difficult to dewater and have a high demand for conditioning chemicals.

A primary objective of conditioning is to increase particle size by combining the small particles into larger aggregates. As biosolids particles are typically negatively charged and repel rather than attract one another, conditioning is used to neutralize the effects of this electrostatic repulsion so that the particles can collide and increase in size (7).

Conditioning is a two-step process consisting of coagulation and flocculation. Coagulation involves destabilization of the biosolids particles by decreasing the magnitude of the repulsive electrostatic interactions between particles. This process occurs through compression of the electrical double layer surrounding each particle.

Flocculation follows coagulation and is the agglomeration of colloidal and finely divided suspended matter by gentle mixing (7).

If the flocculated biosolids are subjected to stress, floc shearing can occur. Therefore, mixing should provide just enough energy to disperse the conditioner throughout the biosolids and bring the particles and colloidal suspensions together. Consideration should be given to providing individual conditioning for each dewatering unit, because it is neither always economical nor good practice to provide one common conditioning unit for several dewatering units. Problems can arise in balancing the flow rates of the various streams when starting up or shutting down individual units. The location of the conditioning unit, relative to each dewatering device, requires optimization.

The amount of conditioning required for biosolids depends on the processing conditions to which the biosolids have been subjected and on the mechanics of the conditioning process available. Both the degree of hydration and fines content of a biosolids stream can be materially increased by exposure to shear, heat, or storage. For example, pipeline transport of biosolids to central processing facilities, weekend storage of biosolids prior to mechanical dewatering, and storage of biosolids for long period of time has been shown to increase the demand for conditioning chemicals prior to all types of dewatering. These factors should be considered in the design of the complete dewatering facility (4).

3. INORGANIC CHEMICAL CONDITIONING

Inorganic chemical conditioning is associated principally with vacuum and pressure filtration dewatering processes. The chemicals normally used in conditioning municipal wastewater biosolids are lime and ferric chloride. Less commonly, ferrous sulfate, ferrous chloride, and aluminum sulfate have been used.

3.1. Ferric Chloride

Ferric chloride is added to biosolids in conjunction with lime and is added first. It hydrolyzes in water, forming positively charged soluble iron complexes that neutralize the negatively charged solids, thus causing them to aggregate. Ferric chloride also reacts with the bicarbonate alkalinity in the biosolids to form hydroxides that act as flocculants. The following equation shows the reaction of ferric chloride with bicarbonate alkalinity:

$$2FeCl_3 + 3Ca(HCO_3)_2 \rightarrow 2Fe(OH)_3 + 3CaCl_2 + 6CO_2 \tag{1}$$

Generally, ferric chloride solutions are used at the concentration received from the supplier (30–40%) because dilution can lead to hydrolysis reactions and the precipitation of ferric hydroxide.

An important consideration in the use of ferric chloride is its corrosive nature. Special materials must be used in its handling, with recommended materials being epoxy, rubber, ceramic, PVC, and vinyl. Contact with skin and eyes must be avoided. Rubber gloves, face shields, goggles, and rubber aprons should be used at all times.

Ferric chloride can be stored for long period of time without deterioration. Usually it is stored in above-ground tanks constructed of resistant plastic or in lined steal tanks. At low temperatures, ferric chloride can crystallize, generally which means that tanks must be stored indoors or must be heated. Table 2 shows the freezing temperature of various concentrations of ferric chloride (4).

Table 2
Crystallization Temperature for Ferric Chloride Solutions

Solution strength (%) $FeCl_3$	Freezing temperature of an unagitated solution °C	°F
20	–21	–5
40	–23	–10
45	–1	+30

Source: US EPA.

3.2. Lime

Hydrated lime is usually used in conjunction with ferric iron salts. Although lime has some slight dehydration effects on colloids, it is chosen for conditioning principally because it provides pH control, odor reduction, and disinfection. $CaCO_3$, formed by the reaction of lime and bicarbonate, provides a granular structure that increases biosolids porosity and reduces biosolids compressibility. Lime is available in two dry forms, quicklime, CaO, and hydrated lime, $Ca(OH)_2$. When using quicklime, it is usually first slurried with water, which converts it to calcium hydroxide prior to adding it to the biosolids. This process, which is called slaking, produces heat and thus special equipment is required. Quicklime normally is available in three grades: high (88–96%) CaO; medium (75–88%) CaO; and low (50–75%) CaO. These grades can affect the slaking ability of the material and should be considered when deciding which grade to purchase. In general, only quicklime that is highly reactive and quick slaking should be used for conditioning. Quicklime must be stored in a dry area, as it reacts with moisture in the air and can become unusable (8).

Hydrated lime is easier to use because it does not require slaking, mixes easily with water (with very little heat produced), and does not require any special storage conditions. However, it is more expensive and less available than quicklime. Thus, the general rule of thumb is to obtain and slake quicklime for applications that require more than 1–2 t/d.

3.3. Dosage Requirements

Iron salts, such as ferric chloride, are usually added at a dose rate of 20–62 kg/t (40–125 lb/t) of dry solids in the biosolids feed, whether or not lime is used. Lime dosage usually varies from 75 to 277 kg/t (150–550 lb/t) of dry solids dewatered. Table 3 lists typical ferric chloride and lime dosages for various biosolids (5).

Inorganic chemical conditioning increases biosolids mass. A designer should expect one pound of additional biosolids for every pound of lime and ferric chloride added (4). This increases the amount of biosolids to be handled and lowers the fuel value for incineration. Nevertheless, use of lime can be beneficial because of its biosolids stabilization effects.

3.4. Other Types of Inorganic Conditioners

Other types of inorganic materials have been used to condition biosolids. The following is a brief description of some of these materials and their uses:

Table 3
Typical Conditioning Dosages of Ferric Chloride and Lime for Municipal Wastewater Biosolids

	Vacuum filter[a] (lb/t)		Pressure filter[a] (lb/t)	
Type of sludge	$FeCl_3$	CaO	$FeCl_3$	CaO
Raw				
Primary	40–80	160–200	80–120	20–280
WAS	120–200	0–320	140–200	400–500
Primary + TF[b]	40–80	180–240	–	–
Primary + WAS[b]	50–120	180–320	–	–
Primary + WAS (septic)	50–80	240–300	–	–
Elutriated aerobically digested				
Primary	50–80	0–100	–	–
Primary + WAS	60–120	0–150	–	–
Aerobically digested				
Primary	60–100	200–260	–	–
Primary + WAS	60–120	300–420	–	–
Primary + TF	80–120	250–350	–	–
Thermally conditioned	None	None	None	None

Source: US EPA.

[a] All values shown are for pounds of either $FeCl_3$ or CaO per ton of dry solids pumped to the dewatering unit. lb/t × 0.5 = kg/T.

[b] TF = trickling filter; WAS waste-activated sludge.

a. **Coal.** Pulverized coal has been used successfully as a conditioning agent in centrifuge and vacuum filter studies done by EPA and others (9). The study by Albertson and Koppers (10) showed that in a concurrent, solid bowl centrifuge, cake solids were increased from 7 to 14% with fine coal addition in the ratio of 0.1:0.3 kg coal/kg dry biosolids (0.1–0.3 lb/lb). The main benefit of fine coal addition centrifuge feed seems to be the improvement in the cake solids concentration. Because of the increased moisture removal provided by the fine coal feed, fuel costs for biosolids combustion can be reduced as much as 60–90%. However, some concerns, including those of safety have arisen concerning materials handling, dust generation, and incinerator temperature control with the addition of coal.

b. **Cement kiln dust.** Cement kiln dust has been used to successfully condition biosolids before dewatering on vacuum filters and also for before and after stabilization. Kiln dust is a byproduct of the cement and lime industries and is high in calcium and potassium. About twice the amount of kiln dust is required to achieve the same pH as from lime. However, the cost is reported to be about 30% that of lime. Some material handing problems have been reported, but the advantages appear to warrant further investigation (11).

c. **Ash.** Flyash, power plant ash, and biosolids incinerator ash can be used as biosolids conditioning agents to increase biosolids dewatering rate, improve cake release, increase cake solids, and in some cases reduce the dosage of other types of conditioning agents. As early as 1927 a process was patented for taking biosolids incinerator ash from an incineration unit back to a vacuum titter to assist dewatering. Ash has been used both as a precoat and as a body feed in one manufacturer's dewatering system with high pressure filtration. However, usually ferric chloride and lime must be used with flyash. The City of Indianapolis, Indiana has successfully used ash as a conditioner on their rotary belt vacuum filters to minimize conditioner requirements and enhance cake release from the media.

4. ORGANIC POLYMERS

During the last two decades, important advances have been made in the manufacture of polymers for use in wastewater biosolids treatment. Polymers are now widely used in biosolids conditioning and a large variety is now available (12–16). It is important to understand that these materials highly differ in chemical composition, functional effectiveness, and cost effectiveness. Reasons for selecting polymers over inorganic chemical conditioners are as follows:

a. Little additional biosolids mass is produced. Inorganic chemical conditioners typically increase biosolids mass by 15–30%.
b. If dewatered biosolids are to be used as a fuel for incineration, polymers do not lower the fuel value.
c. They allow for cleaner material-handling operations.
d. They reduce operation and maintenance problems.

Potassium permanganate, which is often used for odor control, has been shown to reduce polymer doses on mechanical dewatering units. Treatment plant operating personnel, who have optimized the polymer use in conjunction with permanganate use, report a 5–15% reduction of polymer. The subject of organic polymers (polyelectrolytes) is covered fully in the following chapter.

5. SELECTION OF A CONDITIONING CHEMICAL

Many factors go into the selection of the appropriate conditioning chemical to be used at a particular plant. These factors include such considerations as performance, material handling, storage requirements, type of dewatering units, final disposal method and economics. Any evaluation of the conditioning process must take into consideration capital, operating and maintenance costs for the entire system and the impact of sidestreams on other plant processes, the plant effluent, and resultant air quality. Figure 2 shows how the evaluation would look in a quantified flow diagram.

This type of analysis is necessary because conditioning processes differ and, therefore, produce differing consequences for the total system. For example, a plant whose final disposal method is incineration wants the driest cake possible with the least mass and the highest volatile content. Therefore, polymer conditioning is usually the better choice when compared with inorganic chemicals. Polymer conditioning also proves to be the better choice if either storage space is at a minimum or if material handling could be a problem.

Manufacturers' representatives can be of more assistance in evaluating conditioning agents. However, there are several tests, which the designer can perform quickly and inexpensively, that will provide a great deal of information about the conditioner's performance. Such tests can also estimate the quantities of chemicals that will be required (17–19). If biosolids are not available from the plant, the designer could use either pilot-plant biosolids or biosolids from a similar plant. However, once the plant is on-line, the conditioner must be re-evaluated.

5.1. Jar Test

The Jar test is used to screen conditioning agents especially when the designer faces a wide variety of potentially effective products. This test is performed by taking tour to

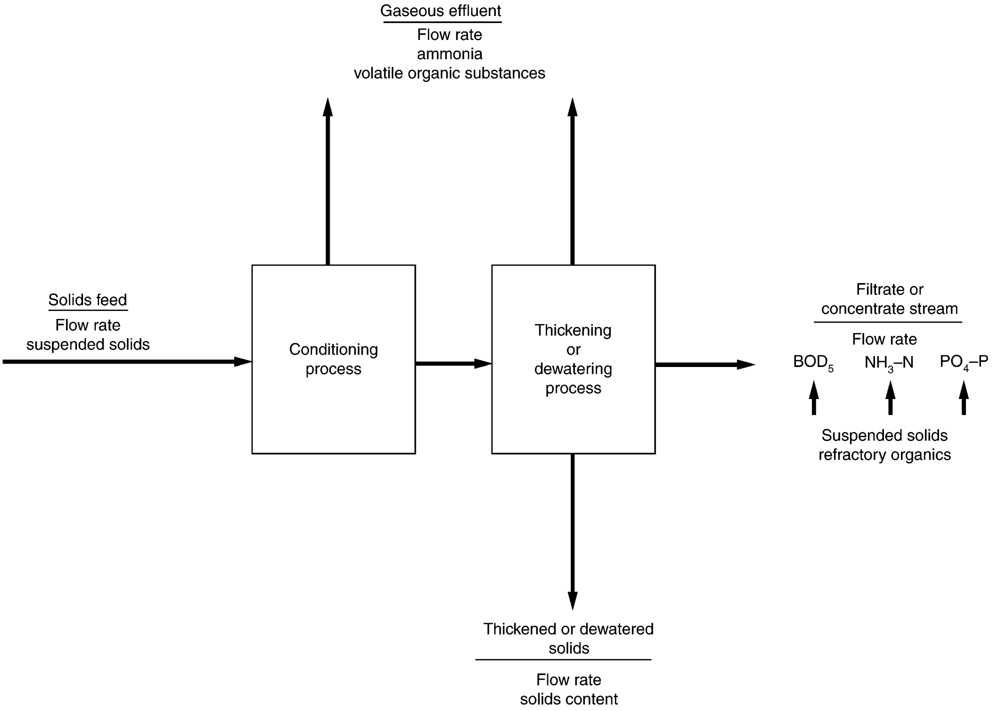

Fig. 2. Basic parameters for evaluation of a biosolids conditioning system (*Source:* US EPA).

six large beakers of about 1-L capacity and filling them with about 600 mL of the biosolids. Solutions of different types of conditioning chemicals are prepared in accordance with the manufacturer's instructions. Conditioning chemicals could include ferric chloride, lime, and up to about four polyelectrolytes. Each of the conditioning chemicals can then be added to a different biosolids sample at the manufacturer's suggested dosage levels or at levels noted previously in this chapter. The beakers are then placed on a gang stirrer with the filter pan in position beneath the oversized paddle. The paddle should just clear the bottom of the pan. The stirrer should be set to 75 rpm. The diluted chemical is then poured into the filter pan and mixed for 30 s. The operator then stops the gang stirrer, removes the paddle, and observes the floc formation and settling (17,20).

5.2. Filter Leaf Testing

The Filter Leaf Test (4,12) is usually used for evaluating dewaterability, primarily by a vacuum filter. Further, in some cases, this test has been used to size a vacuum filter. The test is performed by assembling a filter leaf apparatus as shown in Fig. 3. The filter cloth should be the fabric intended for use or monofilament filter.

A jar test apparatus is used to prepare chemically conditioned biosolids in at least 2-L batches for each filtration cycle. The conditioning chemicals are placed in the jar test apparatus, allowing for 2–4 min of mixing and flocculation time. The mixing should be slow (about 10 rpm). Flash mixing will adversely affect test results.

Two liters of the chemically conditioned biosolids are transferred to a beaker, its temperature is measured and the filter leaf is submerged in it about 5 cm (2 in.) below the

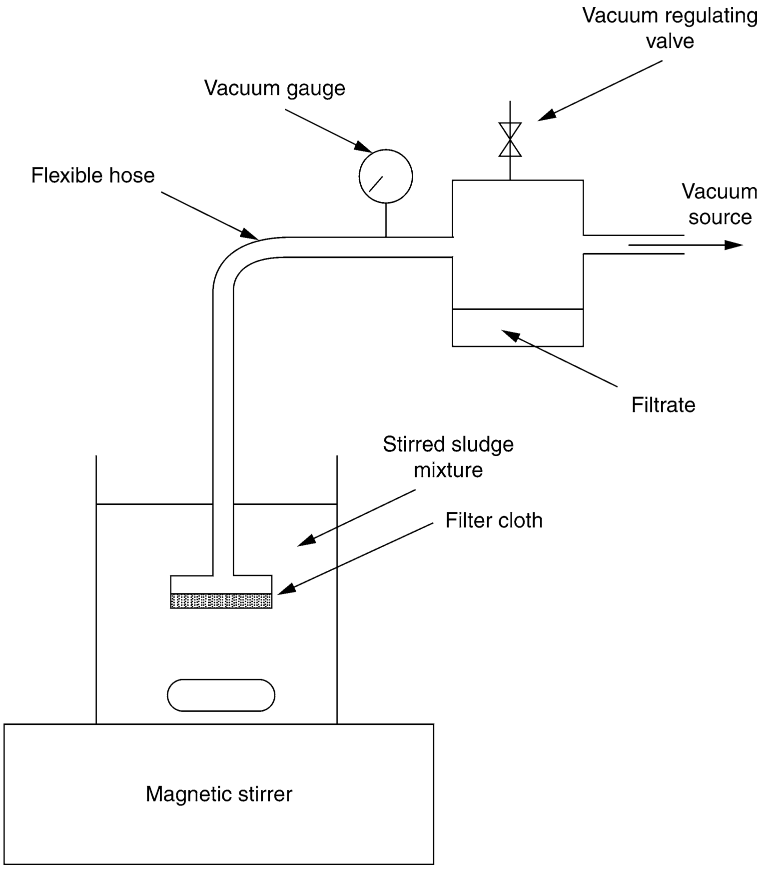

Fig. 3. Filter leaf apparatus (*Source:* US EPA).

surface. Then 51 cm (20 in.) Hg of vacuum is applied to the filter leaf and timing begins. After 45 s of form time, the leaf is withdrawn and dried for 90 s.

The cake thickness is measured and the cake is scraped into a previously weighed dish. The dish and cake are weighed and transferred to a drying oven. After air drying, the cake should be desiccated, weighed, volatilized, desiccated again, and weighed again.

The following determinations should be made for each run:

a. Volume of filtrate (mL).
b. Temperature of filtrate (°C).
c. Wet weight of filter cake (g).
d. Dry weight of filter cake (g).
e. Dry weight of ash (g).
f. Total solids concentration of cake (Ts) (wet weight [%]).
g. Volatile solids concentration of cake solids (Vs) (total solids [%]).

This test simulates a 3-min cycle time, divided into 45 s of form time, 90 s of drying time, and 45 s of discharge time. It may be desirable to use a longer cycle time. Typically, cake thickness with cycle time and cake total solids content will increase. Cake solids content can be further increased by decreasing the ratio of form time to drying time. Subsequently, experiments might be conducted to determine yield as a

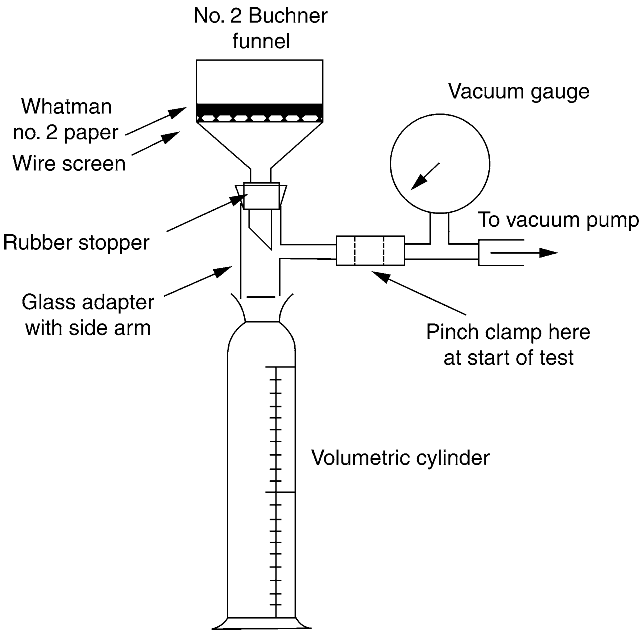

Fig. 4. Buchner funnel apparatus (*Source:* US EPA).

function of percent solids concentration and/or a function of cycle time. Also for each run the filter yield, in lb cake solids/ft^2 of filter/h can be calculated with Eq. (2):

$$\text{Filter yield} = \text{Dry weight of cake (g)} \times \text{cycles/h} + 453.6 \times \text{filter area } (\text{ft}^2) \qquad (2)$$

One of the advantages of the filter leaf test is that it simulates actual behavior on a vacuum filter. Ease of cake release from the filter cloth can be estimated, and the percent moisture of the final cake can be determined. The actual sizing of a vacuum filter would require a series of leaf tests, using one or more controls in which no characteristics are added. A range of chemical dosages and combinations should then be studied.

5.3. Buchner Funnel Test for Determination of Specific Resistances

The specific resistance test is another method of predicting conditioning agent performance (21,22). The Buchner funnel test equipment (19) consists of a graduated cylinder, Buchner funnel, and a vacuum pump as shown in Fig. 4.

A series of conditioned biosolids samples are prepared in large beakers as previously discussed in Sections 5.1 and 5.2. First about 200 mL of the thickened biosolids is placed into the beakers. The biosolids tested should be representative of the biosolids to be used on the dewatering units. These biosolids can be from a pilot-plant or a similar full-scale treatment plant (19).

A Buchner funnel is mounted on top of a graduated cylinder as shown in Fig. 3, and the funnel is fitted with a piece of filter paper. For each test, a portion of the conditioned biosolids (50–200-mL) is poured into the funnel. After 2 min of gravity drainage, the vacuum pump is turned on (15 in Hg). At about 15-s intervals, the filtrate volume is measured and recorded until the vacuum breaks or additional water can not be removed.

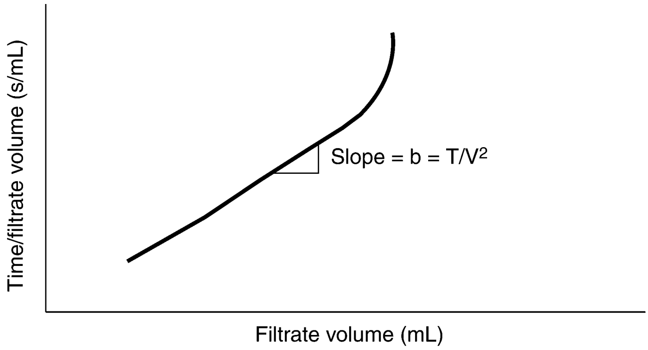

Fig. 5. Time/filtrate volume vs filtrate volume plot (*Source:* US EPA).

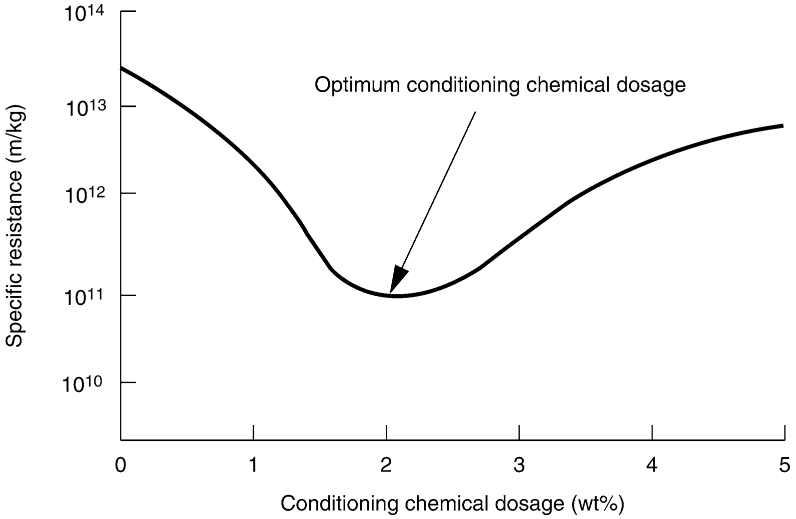

Fig. 6. Plot of specific resistance vs conditioning chemical dosage (*Source:* US EPA).

The biosolids cake is then removed from the filter and placed in a weighed dish. The wet weight of the cake is measured and then after drying at 180°C, the dry weight is measured. Total suspended solids value is determined on the filtrate sample. In addition, the temperature of the filtrate is also measured. A plot is made of time/filtrate volume versus filtrate volume, as shown in Fig. 5. The slope of the straight line portion of the graph is "b" and is used to calculate the specific resistance (r) from Eq. (3).

$$r = (2PA^2b)/\mu w \tag{3}$$

where r is the specific resistance (m/kg); P is the pressure of filtration (N/m^2); A is the area of filter (m^2); b is the slope of time/volume vs volume curve (s/cm^6); μ is the viscosity of filtrate (N s/m^2); and w is the weight of dry solids/volume of filtrate (kg/m^3).

Figure 6 shows a plot of specific resistance vs conditioning chemical dose. This plot was constructed with specific resistance data from biosolids conditioned with different levels of the same chemical. From a plot such as this, the designer can determine optimum

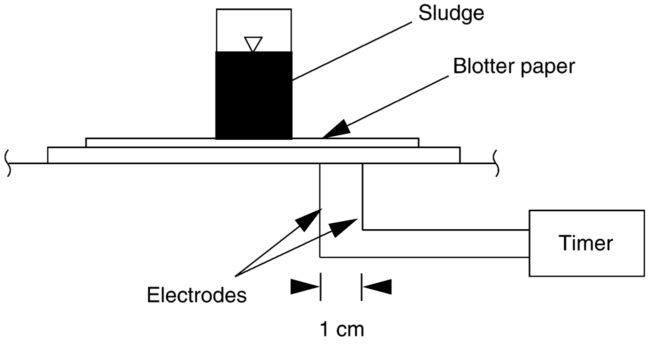

Fig. 7. Capillary suction time apparatus (*Source:* US EPA).

chemical dose. The optimum conditioner chemical dosage is that which produces the lowest specific resistance.

A modification of the Buchner funnel test can be used to duplicate the gravity drainage results which can be achieved on a belt filter press. This test uses the apparatus shown in Fig. 3, exclusive of the vacuum pump. A piece of the belt material which will be used in the filter press is placed in the Buchner Funnel. A sample of conditioned biosolids is placed into the funnel and the volume of water released is measured at regular intervals. Both time and volume are recorded. The polymer and/or dose of polymer which gives the greatest volume of free water in the shortest time should give the best results on the belt filter press.

5.4. Capillary Suction Time

The capillary suction time (OST) is a simple and quick test that measures the time required for the liquid portion of the biosolids to travel 1 cm or any other fixed distance (23). The apparatus (Fig. 7) consists of a timing device, an upper plate containing probes that activate and deactivate the timing device, and a lower plate that holds the filter paper and a metal sample container.

A sample of conditioned biosolids is placed in the sample container. As water migrates through the paper and roaches the first probe, it activates the timer. When the water reaches the second probe, the timer deactivates. The time interval between timer activation and deactivation is the OST and is a measure of the dewaterability of the conditioned biosolids. OST is plotted vs chemical dosage. The dosage that gives the fastest time is the optimum. Conditioner types and concentrations should be varied until the optimum chemical and dosage is found for a particular dewatering system.

6. COST

6.1. Capital Cost

Figure 8 shows the relationship between construction costs of ferric chloride storage and feed facilities and installed capacity. For example, if a designer needed to feed 100 lb/h (45.4 kg/h) of ferric chloride the estimated cost would be 330,000 USD. All costs are given in 1975 dollars (Cost Index = 190.49). To obtain the values in terms of the present 2006 USD using the Cost Index for Utilities shown in Appendix (24), multiply the costs by

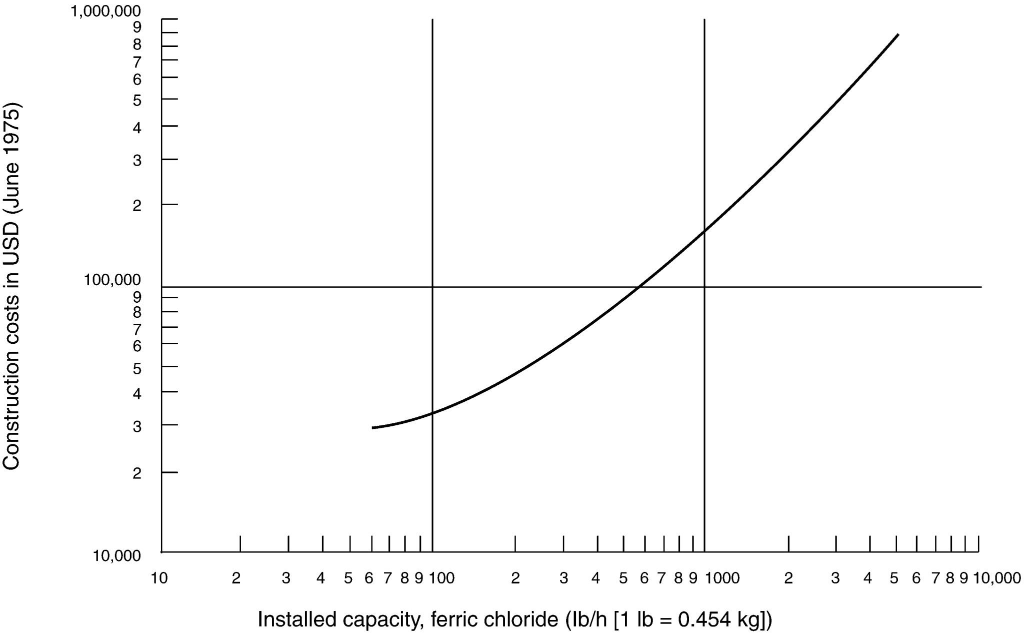

Fig. 8. Capital costs of ferric chloride storage and feeding facilities (*Source:* US EPA).

a factor of 2.77. Costs for Fig. 8 are estimated on the basis of liquid ferric chloride use. Chemical feed equipment was sized for a peak feed rate of twice the average. At least 15 d of storage was provided at the average feed rate. Piping and buildings provided to house the feeding equipment are included.

Figure 9 gives construction costs of lime storage and feeding facilities as a function of installed capacity. Cost estimates shown on Fig. 9 are based on the use of hydrated lime in small plants, 50 lb/h (22.7 kg/h) or less, and pebble quicklime in larger plants. Allowances for peak rates of twice the average are built into the lime feed rates. At least 15 d of storage is provided for at the average rate. Storage time varies from installation to installation because it is dependent upon the relative distance to and reliability of the chemical supply. Piping and buildings to house the feeding equipments are included in the estimates. Estimated costs of steel bins with dust collector vents and filling accessories are also included.

6.2. Operation and Maintenance Cost

Figure 10 indicates the relationship between man-hours spent annually for operation and maintenance and lb $FeC1_3$ fed/h. The labor includes unloading the ferric chloride and the operation and maintenance of the chemical feed equipment. Unloading requirements are as follows: for a 4000-gal (15.1 m^3) truck is 1.5 man-h and for 50-gal (0.19 m^3) barrels, 72/truck is 9 man-h. These requirements are shown as man-h/lb of chemicals fed to the process. Metering pump operations and maintenance are estimated at 5 min/pump/shift.

Figure 11 indicates annual electric power requirements for a ferric chloride chemical feed system.

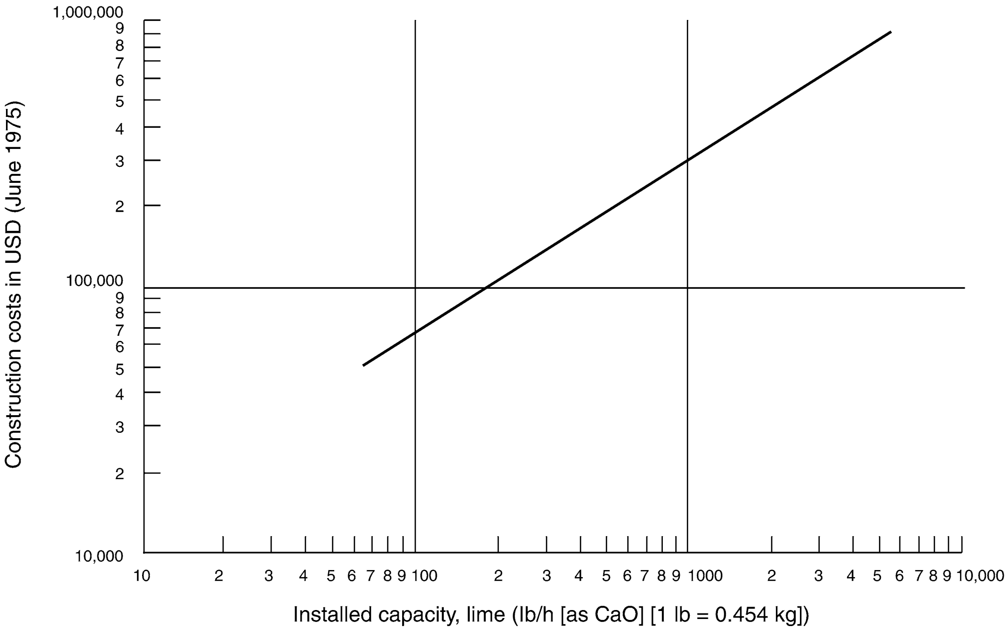

Fig. 9. Capital costs of lime storage and feeding facilities (*Source:* US EPA).

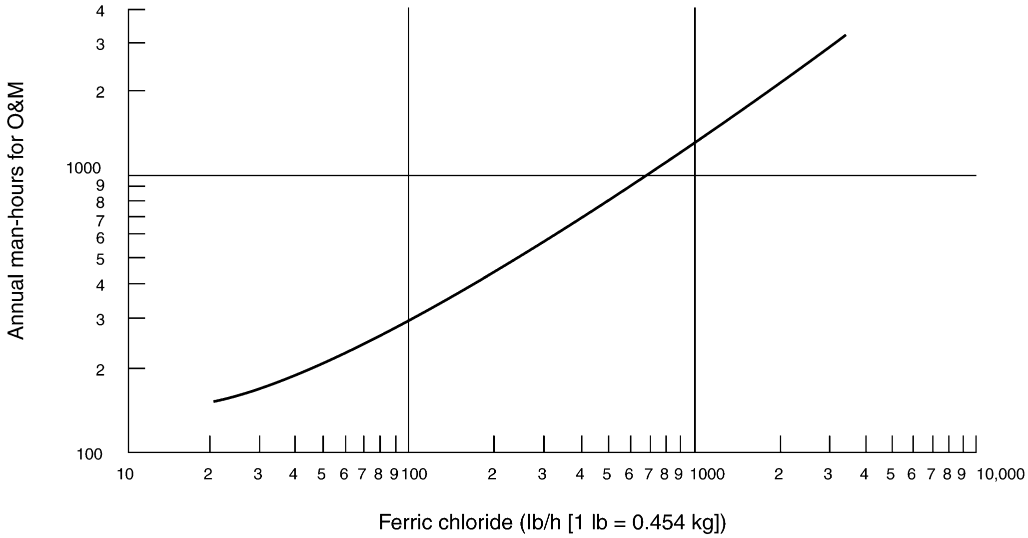

Fig. 10. Ferric chloride storage and feeding O&M work-hour requirements (*Source:* US EPA).

Annual maintenance material costs are typically 3–5% of the total chemical feed system equipment cost. Figure 12 indicates man-hour for operation and maintenance as a function of lb lime fed/h. The curve consists of lime unloading requirements and labor related to operation and maintenance of the slaking and feeding equipment. These requirements are summarized as follows:

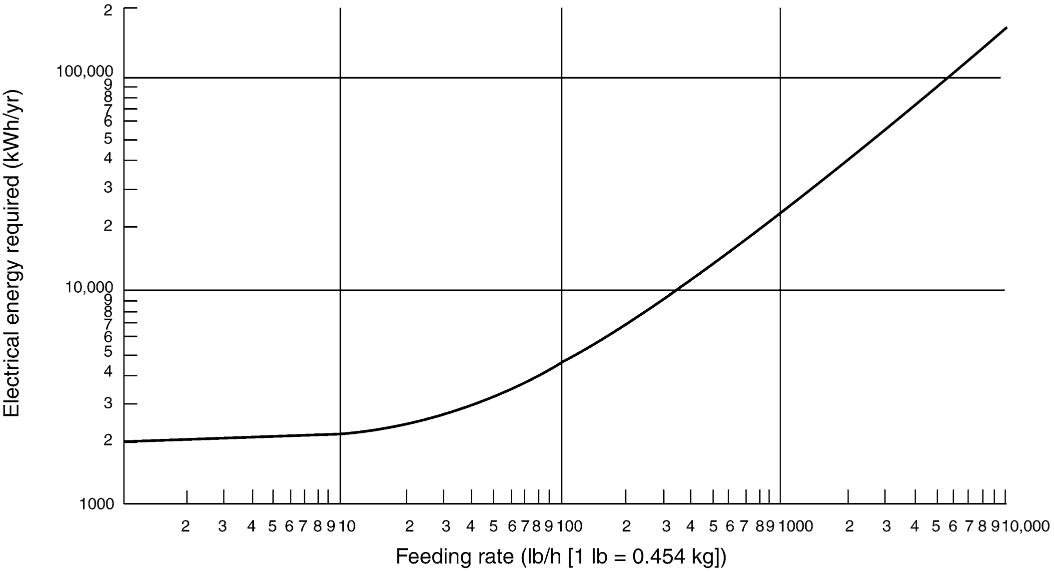

Fig. 11. Electrical energy requirements for a ferric chloride chemical feed system (*Source:* US EPA).

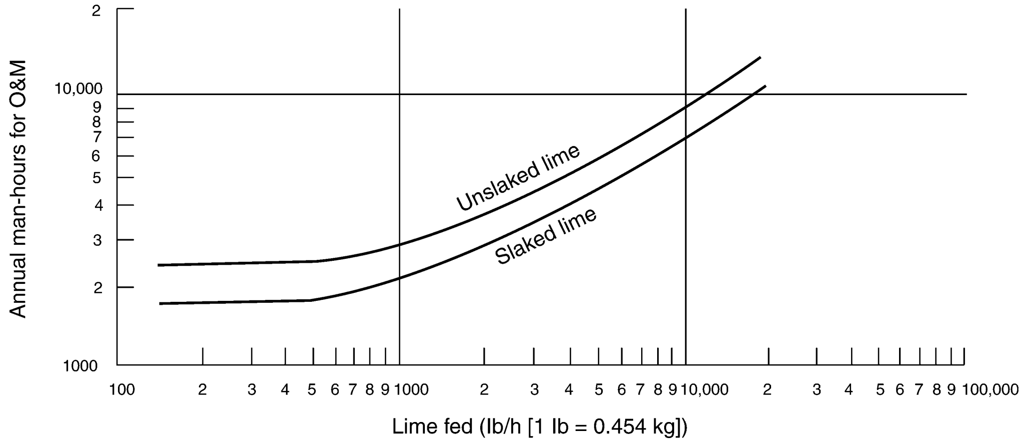

Fig. 12. Lime storage and feeding OM work-hour requirements (*Source:* US EPA).

a. **Slaker.** 1 h/8-h shift/slaker in use.
b. **Feeder.** 10 min/h/feeder.
c. **Slurry pot feed line (for slaked lime).** 4 h/wk.

Figure 13 shows annual electric power requirements for a lime feed system. The major components and the values used in the curves, all expressed kW/h/1000 lb (454 kg) of lime fed are as follows:

a. **Slakers.** 1.6–0.8.
b. **Bin activators.** 2.7–0.36.
c. **Grit conveyors.** 0.45–0.06.
d. **Dust collection fans.** 0.04–0.02.

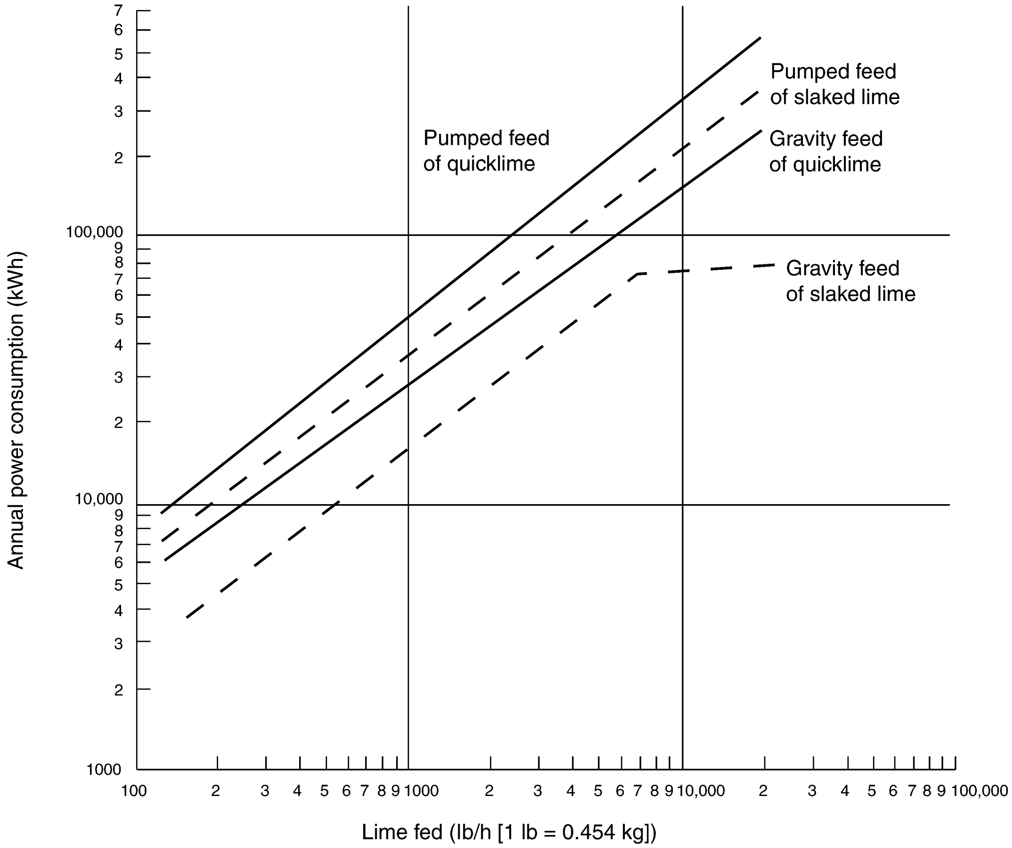

Fig. 13. Electrical energy requirements for a lime feed system (*Source:* US EPA).

e. **Slurry mixers.** 0.027–0.020.
f. **Slurry feed pumps.** 2.2–1.4.

Annual maintenance material costs are typically 0.5–1.5% of the total lime feed system equipment cost.

7. THERMAL CONDITIONING

The thermal conditioning process enhances the dewatering characteristics of biosolids through the simultaneous application of heat and pressure. It is a continuous flow process in which biosolids are heated to temperatures of 177–204°C (350–400°F) in a reactor under pressure of 1720–2750 kPa (250–400 psig) for 15–40 min (25). There are two basic modifications of the thermal conditioning process used in wastewater treatment. In one modification, low pressure oxidation (LPO), air is added to the process. The other modification, heat treatment (HT), does not include the addition of air to the process. Both thermal conditioning processes produce biologically stable biosolids with excellent dewatering characteristics.

Wastewater biosolids contain water and cellular and inert solids that form a gel-like structure. The water portion consists of bound water, which surrounds each solids particle, and water of hydration, which is inside the cellular solids. Thermal conditioning improves biosolids dewaterability by subjecting the biosolids to elevated temperature and pressure in a confined reactor vessel: thus coagulating solids, breaking down the gel-like structure of the biosolids, and allowing the bound water to separate from the solids particles. In addition, hydrolysis of protein material in the biosolids occurs. Cells break down and water is released, resulting in coalescence of solids particles. In its conditioned state, the biosolids are readily dewatered on most dewatering devices to 30–50% solids, in most cases without addition of chemicals.

A portion of the volatile suspended solids (VSS) in biosolids is solubilized as a result of the breakdown of the biosolids structure. The solubilization of VSS increases its biodegradability. Although this solubilization does not change the total organic carbon content of the biosolids, it does result in an increase in the BOD_5. The BOD_5 produced is of primary concern in the recycle of sidestreams. The solubilization of VSS and the resultant BOD_5 production for HT systems may be estimated (25) with Eq. (4):

$$\mathrm{VSS} = 0.1\ \mathrm{PB} + 0.4\ \mathrm{WAS} \tag{4}$$

$$\mathrm{BOD_5} = 0.07\ \mathrm{PB} + 0.3\ \mathrm{WAS}$$

where VSS is the volatile suspended solids solubilized (dry kg [lb]); PB is the primary biosolids (dry kg [lb]); WAS is the waste-activated sludge (secondary biosolids) (dry kg [lb]); and BOD_5 is the 5-d biochemical oxygen demand produced by VSS solubilization (kg [lb]).

Using these rule-of-thumb procedures, 9.9 kg (22 lb) of VSS solubilization and 7.3 kg (16 lb) of BOD_5 are produced by HT of 45kg (100 lb) of a typical mixture of 60% primary biosolids and 40% WAS. In LPO systems, VSS solubilization and BOD_5 production are expected to be approximately the same.

Thermally conditioned biosolids can be dewatered on vacuum filters, belt filter presses, recessed plate filter presses, centrifuges, or sand drying beds. The dewatered solids can then be incinerated, disposed of in a landfill or recycled by land application.

7.1. HT Process

A schematic diagram of a typical HT system is shown in Fig. 14. In this continuous process, biosolids are ground to reduce particle size to less than 0.64 cm (0.25 in.) and are then pumped through a heat exchanger and into a reactor. Normal discharge pressure from the biosolids feed pump is approx 1720 kPa (250 psi). In the heat exchanger, the temperature of the biosolids is raised from ambient to between 149 and 177°C (300 and 350°F). The heated biosolids exit the heat exchanger and enter a reactor feed standpipe, where steam is injected through a nozzle and the biosolids are mixed turbulently. The steam and biosolids proceed upward through the standpipe and enter the reactor at the top. The hot biosolids are retained for a period of time in the reactor and are subsequently returned through the heat exchanger to be cooled at approx 49°C (120°F). From the discharge side of the heat exchanger, the conditioned biosolids flow through a control valve, which controls reactor biosolids level and pressure, and into a decant tank. The decant tank permits rapid settling and compaction of the biosolids particles and the release of gas. The settled biosolids are pumped to a dewatering device. Process off-gases can be treated by various odor control methods (25,26).

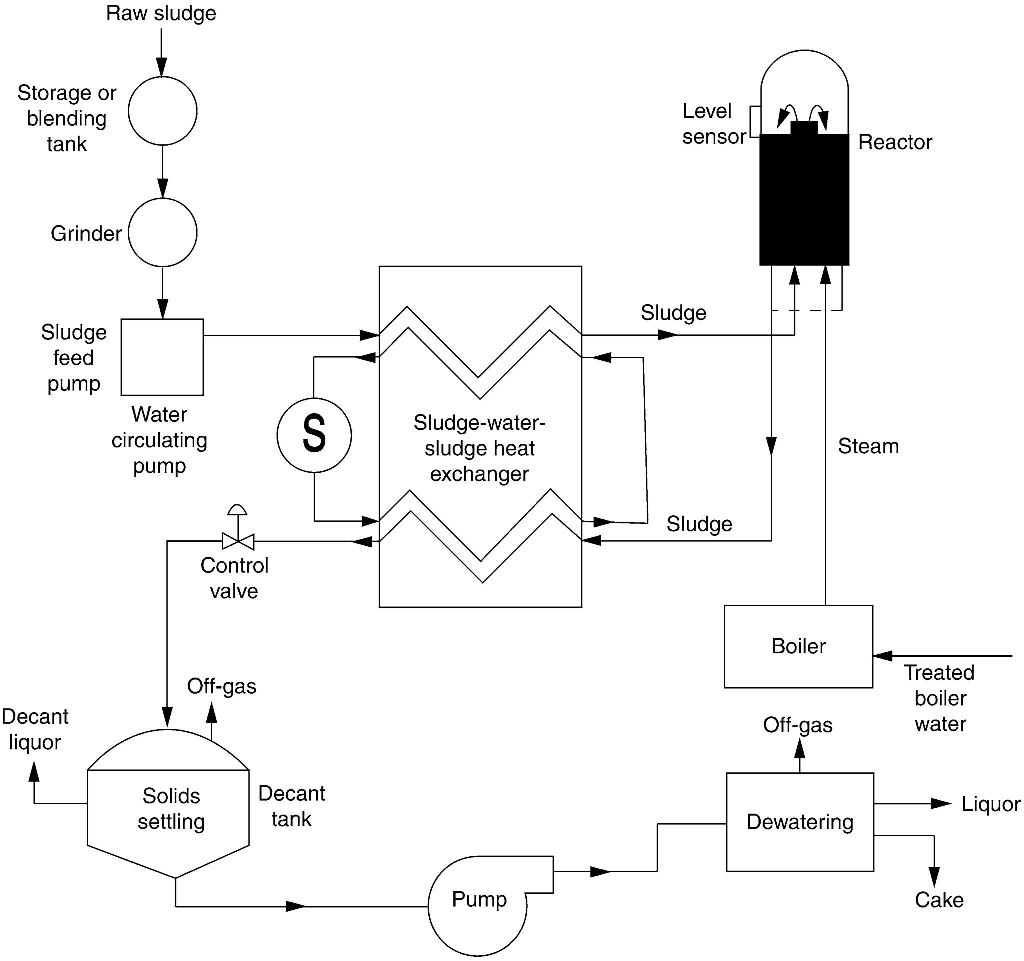

Fig. 14. Heat treatment (HT) process flow diagram (*Source:* US EPA).

7.2. LPO Process

A schematic diagram of the LPO system is shown in Fig. 15. Raw biosolids are first passed through a grinder in which particles are reduced to less than 0.64 cm (0.25 in.) in size. The ground biosolids are then pumped at approx 2750 kPa (400 psi) through a heat exchanger followed by an LPO reactor, High pressure air from the system air compressor is introduced into the biosolids flow upstream of the heat exchanger. The air improves heat transfer and converts sulfur products in the biosolids to sulfate, slightly reducing odor from off-gases. The resulting turbulent flow of biosolids and air proceeds through the heat exchanger where biosolids are preheated by processed biosolids returning from the LPO reactor. The biosolids and air mixture enters the reactor at a temperature between 149 and 160°C (300 and 320°F). Steam is injected directly into the reactor to increase the biosolids/air mixture temperature to between 166 and 177°C (330 and 350°F). The combined products rise slowly in the reactor and a slight heat of reaction or oxidation occurs, producing a small

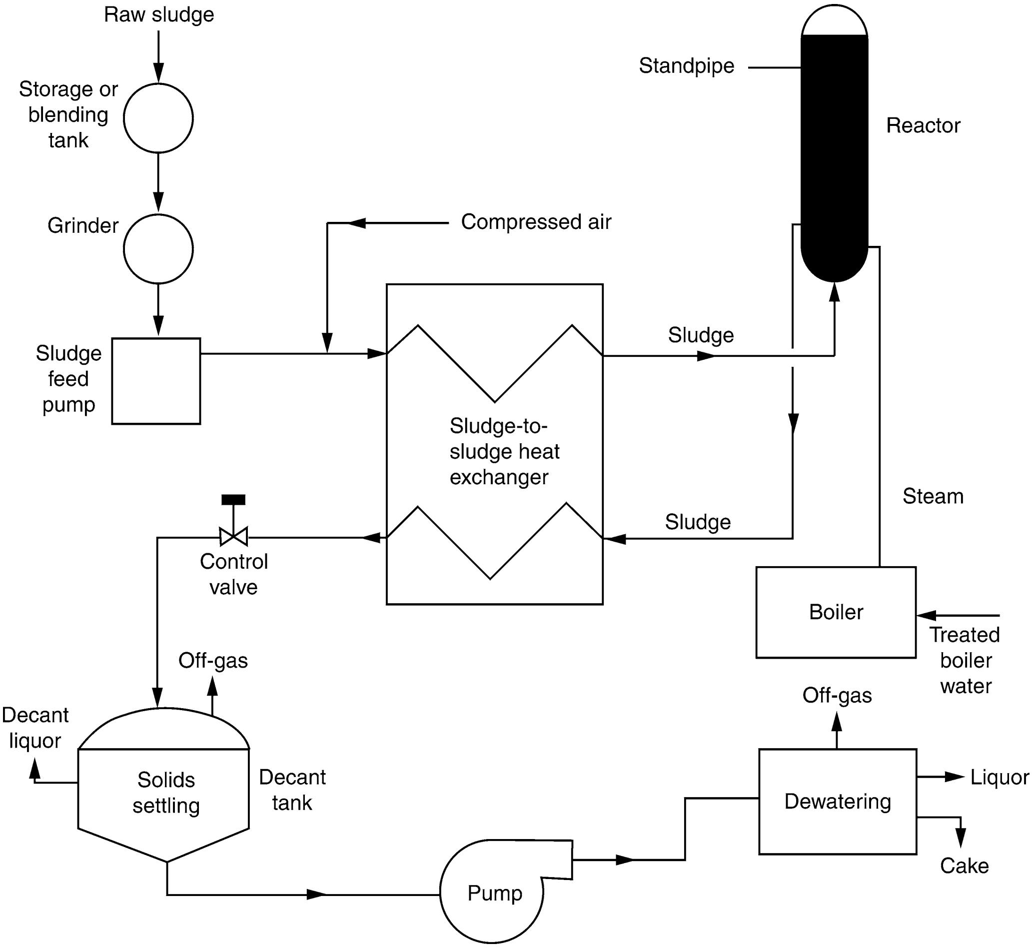

Fig. 15. Low pressure oxidation (LPO) process flow diagram (*Source:* US EPA).

amount of heat. From the reactor midpoint to the reactor outlet, the biosolids temperature increases approx 10°C because of the heat of reaction of the biosolids, contributing to an overall temperature increase from the reactor inlet to reactor outlet of approx 40°C. Detention time or "cook time" in the reactor is based on the volume of the reactor and the height of the discharge pipe (standpipe or downcomer line). The detention time is controlled by the air, steam, and biosolids flow rates to the reactor (25,26).

After leaving the LPO reactor, the partially oxidized product flows back through the heat exchanger and releases heat to the incoming biosolids/air mixture. When the partially oxidized product reaches the control valve, the temperature ranges between 43°C and 54°C (110°F and 130°F). This valve controls the pressure in the reactor. From the valve, the thermally conditioned biosolids and exhaust gases are released. The settled solids are then pumped to a dewatering device before final disposal. Process off-gases from the LPO system also can be treated by various odor control methods.

7.3. Economic Considerations

Since the early 1970s, the increase in the cost of natural gas and fuel oil has significantly changed the economic feasibility of new thermal conditioning systems for small plants. Larger installations more than 0.44 m^3/s (10 MGD) that use dewatering and incineration with energy recovery may determine that the addition of a thermal conditioning step would be an economic asset.

Several factors must be considered regarding the cost effectiveness of a thermal conditioning system as a function of plant size (25).

a. Present-day energy costs dictate some form of heat recovery to make the thermal conditioning process competitive with other conditioning processes.
b. Thermal conditioning systems require well trained and skilled supervisors and operators for optimization of the operation and maintenance of the systems.
c. Both types of systems should be supported with a complete inventory of spare parts to reduce excessive downtime. Also, they require a thorough preventive maintenance program.
d. The unit capital cost of thermal conditioning systems in 2004 USD is in the range of 562–803 USD/MG (510–730 USD/t) of annual biosolids production, when processing over 9090 dry ton (10,000 t)/yr because of the use of multiple treatment units and standby units rather than larger sized individual units. At lower loading rates, processing costs increase significantly, and the comparatively high cost of support systems (such as boilers, air compressors, and decant tanks) makes HT/LPO systems more costly to build than other biosolids conditioning facilities.

7.4. Advantages and Disadvantages of HT/LPO Conditioning

Previous literature (5,25,26) on HT/LPO provides a summary of the advantages and disadvantages of using these processes to condition wastewater biosolids.

Advantages are as follows:

a. Except for straight WAS, the process produces biosolids with excellent dewatering characteristics. Cake solids concentrations of 30–50% can be obtained with conventional dewatering equipment.
b. The processed biosolids do not normally require chemical conditioning to dewater well on mechanical equipment.
c. The process stabilizes and sterilizes the biosolids by destroying all living organisms including pathogens.
d. The process provides biosolids with a heating value of 26,000–30,000 kJ/kg (11,000–13,000 Btu/lb) of volatile solids, suitable for incineration or anaerobic digestion with energy recovery.
e. The process is suitable for many types of biosolids that cannot be stabilized biologically because of the presence of toxic materials.
f. The process is effective on feed biosolids with a broad range of characteristics and is relatively insensitive to changes in biosolids characteristics.
g. Continuous operation is not required as with incineration, as the system can easily be placed on standby.

The disadvantages are as follows:

a. The process has high capital costs owing to mechanical complexity and the use of corrosion-resistant materials, such as stainless steel, in the heat exchangers.
b. The process requires careful supervision, skilled operators, and a good preventive maintenance program.

c. The process produces a malodorous gas stream that must be collected and treated before release.
d. The process produces dark colored sidestreams with high concentrations of organics and ammonia nitrogen.
e. Scale formation in heat exchangers, pipes, and the reactor requires cleaning by difficult and/or hazardous procedures.
f. Subsequent centrifugal dewatering may require continuous or intermittent polymer dosage to control recycle of fine particles.
g. The daily biosolids throughput of the process cannot be adjusted by a significant amount without incurring high energy and/or labor costs.

For further information and details on thermal conditioning, the reader is referred to refs. 26–33.

8. MISCELLANEOUS CONDITIONING PROCESSES

8.1. Elutriation

Elutriation is the term commonly used to refer to the washing of anaerobically digested biosolids before vacuum filtration. Washing causes a dilution of the bicarbonate alkalinity in the biosolids and therefore reduces the demand for acidic metal salt by as much as 50% (34). Full discussion on elutriation is presented in the next chapter.

The process itself was patented by Genter in 1941 (35). Although it typically uses one or two tanks, any number of tanks can be used. Two to six volumes of washwater, typically plant effluent, flow countercurrent to one volume of anaerobically digested biosolids. Elutriation tanks are designed to act as gravity thickeners, with a mass solids loading of 8–10 lb/ft^2/d (39–49 kg/m^2/d).

Currently the process is not used as extensively as it had been because, in addition to reducing alkalinity, it also washed out 10–45% of the solids from the incoming biosolids stream (36–40). Elutriate was recycled back to the main plant and eventually degraded the plant effluent (37,38,40).

Full-scale research (40–42) has shown that the solids problem can be solved, and 90–92% capture achieved, with the use of polymers. Recommended current elutriation design considerations are listed as follows:

a. Tanks should be loaded at hydraulic loadings (total of both biosolids and washwater flow) of 200–300 gal/d/ft^2 (69–104 L/d/m^2) and solids loading of 8–15 lb/d/ft^2 (39–73 kg/d/m^2).
b. Tanks should have the best possible inlet structure to minimize inlet momentum.
c. Baffling should be used to prevent tank currents.
d. Tanks should be provided with scum collection.
e. Polymer addition should be provided.

8.2. Freeze–Thaw

In 1929, Babbit and Schlenz demonstrated the benefit of freezing wastewater biosolids (43). They noted that, after biosolids were frozen on a sand drying bed during the winter and thawed in the spring, its drainage qualities were improved and it dried to higher solids content. Research has since been conducted in three areas of freeze conditioning: indirect and direct mechanical systems and natural freezing. Most mechanical freeze-conditioning research has been oriented toward indirect freezing methods.

Table 4
Comparison of Biosolids Handling and Conditioning Processes

Process	Reduction in sludge COD (%)	Sludge solubilization	Supernatant and filtrate quality	
			pH	Quality
Slurry freezing	35	Low	7.0–8.0	Good
Solid freezing	50–70	High	7.0–7.5	Poor
Anaerobic digestion	60–70	High	6.0–7.0	Poor
Aerobic digestion	30–70	Low	4.0–7.0	Good
Chemical addition	20–40	Low	6.0–6.5	Moderate

Source: US EPA.

Indirect freezing involves the separation of the refrigerant and the biosolids by some type of partition. The studies on wastewater biosolids indicate that freezing (44–47):

a. Causes cellular dehydration and thus allows better flocculation.
b. Destroys the sliminess of biological biosolids.
c. Improves dewatering characteristics as measured by sand bed and vacuum filter dewatering rates.

Although freeze conditioning has been shown to be beneficial, it is expensive to implement. This is because the system cannot utilize the heat generated by the fusion of the frozen biosolids to cool the refrigerant. To overcome this problem, pilot work has been conducted on direct freezing (48). In direct freezing, the liquefied refrigerant is vaporized and dispersed through the biosolids slurry at a controlled rate. In Table 4, slurry freezing (direct mechanical method) is compared to solid freezing (indirect freezing) and several other treatment processes.

In natural freezing, the freezing is done by the environment. At least one facility (49) is operating in Canada, and extensive full-scale research is being conducted in facility design in order to improve this method of conditioning (50).

8.3. Mechanical Screening and Grinding

In some applications, screening or grinding can be considered part of the biosolids conditioning process. A good example of screening for conditioning is in the application of a disc nozzle centrifuge. A stainless steel, self-cleaning screen is required to remove large solids and fibrous material that would clog the disc nozzle machine. Grinding of primary biosolids is an important step for some biosolids handling processes. It has also been indicated that grinding of thick (>8% solids) biosolids streams reduces viscosity, thus making the slurry easier to pump. One outstanding example of this is in the municipal system at Glen Cove, New York. In addition to the more commonly known conditioning methods previously discussed, research has also been conducted on more novel methods, such as bacteria, electricity, solvent extraction, and ultrasonic.

8.4. Bacteria

Autotrophic sulfur bacteria may provide conditioning if added to digested biosolids before dewatering (34). Under aerobic conditions, sulfur-oxidizing bacteria stimulate the production of sulfuric acid, which, in turn, lowers the pH of the biosolids and enhances the dewatering process as measured by the specific resistance test. In another study (51), it was shown that filtration rates of WAS could be increased under anaerobic conditions with the use of the enzyme lysozyme.

8.5. Electricity

In extensive laboratory and pilot plant work studies, graphite anodes and iron cathodes have been used to condition biosolids (52–58). These studies indicate that:

a. At pH values lower than four electrical current can condition biosolids for filtration without the use of chemicals.
b. The quantity of water removed during dewatering (vacuum filtration) was proportional to the amount of electricity used. Thinner biosolids required less current.
c. Biosolids electrically conditioned seemed to produce drier cakes than chemically conditioned biosolids.

The disadvantages are as follows:

a. Anodes had to be replaced frequently because a dried crust continually formed on them.
b. The system uses a great deal of electricity; optimum current density was approx 0.3 amp/ft^2 (3.3 amp/m^2) of anode surface, with a potential drop of 4 V between the electrodes.
c. No full-scale facilities have been tested to evaluate operating problems.

8.6. Solvent Extraction

Solvent extraction is covered in another book of this series (59) and the principles and practices of solvent extraction are very well explained elsewhere (59–61). In 1957, research was conducted at Rockford, Illinois, with carbon tetrachlorethylene as the solvent, with distillation end products being dried oils, fats, and greases (38). It was not considered to be very economical at that time. Although solvent extraction is becoming popular in industry (62,63), no municipal installations are known to be using the process.

8.7. Ultrasonic

Conditioning of municipal biosolids by ultra- or supersonic vibration has been explored (34). Ultrasonic vibrations degasify biosolids, which is beneficial, but the vibrations also tend to destroy biosolids flocs, resulting in fine solids that are difficult to dewater.

9. BIOSOLIDS STABILIZATION

The principal purposes of stabilization are to make the treated biosolids less odorous and putrescible and to reduce the pathogenic organism content. Some procedures used to accomplish these objectives can also result in other basic changes in the biosolids. The selection of a certain method hinges primarily on the final disposal/reuse procedure planned. If the biosolids are to be dewatered and incinerated, frequently no stabilization procedure is used. Most stabilization methods, particularly anaerobic and aerobic digestion, result in a substantial decrease in the amount of suspended solids present. Hence, the corollary function of conversion is included in the description of these processes.

This chapter provides detailed discussion of one process, chlorine oxidation, which has the primary function of biosolids stabilization. The other stabilization processes namely, anaerobic digestion, aerobic digestion and lime stabilization are covered at length in other chapters of this book. Both anaerobic and aerobic digestion are currently increasing in popularity. The former is receiving revived attention from some cities and new attention from others for several reasons. The production of methane in anaerobic digestion is attractive in view of energy shortages, as is the suitability of anaerobically digested biosolids to disposal on land. Also, it is being recognized that problems experienced previously with anaerobic digestion were actually because of other wastewater process considerations. Interest in aerobic digestion of excess activated sludge is growing because it has the potential for providing a good quality liquid process stream and can produce exothermic reaction conditions. A major impetus for processes such as anaerobic and aerobic digestion and lime treatment is the growing emphasis on utilization of biosolids rather than mere disposal. Chlorine oxidation is of limited use for special situations or where septic tank wastes are involved (5).

Several other biosolids treatment processes provide varying degrees of stabilization, although this is not their principal function. Composting is practised in several US cities and is being actively investigated for others. This process is considered important enough, with the emphasis on recycling of biosolids to the land, that it is discussed in a separate chapter. Heat treatment, discussed earlier, has been installed in several new United States plants to improve biosolids conditioning and dewatering economics. Some processes used to disinfect biosolids, such as heat drying and pasteurization, also provide limited stabilization.

10. CHLORINE STABILIZATION

Stabilization by chlorine addition was developed as a proprietary process and is marketed under the registered trademark “Purifax.” The chlorine stabilization process is applied to wastewater treatment plant biosolids and sidestreams to reduce putrescibility and pathogen concentration. The process has also been used to improve the dewaterability of digested biosolids and to reduce the impact of recycled digester supernatant on the wastewater treatment systems. Because chlorine reactions with biosolids are very rapid, reactor volumes are relatively small resulting in reduced system size and initial costs. The process results in no appreciable destruction of volatile solids, and unlike anaerobic digestion, yields no methane gas for energy generation and little biosolids mass reduction.

Chlorine-stabilized biosolids are buff-colored, weak in odor, sterile, and generally easy to dewater, either mechanically or on drying beds. The stabilized biosolids have been used as soil conditioner. However, there is concern about its use on cropland and its disposal in landfills because of its high acidity, high chloride content, and potential for releasing chlorinated hydrocarbons and heavy metals (5). The stabilized biosolids are corrosive unless pH has been adjusted. Process equipment that comes into contact with biosolids that have not been neutralized must be constructed of acid-resistant materials or be coated with protective films.

10.1. Process Description

Chlorine treatment stabilizes biosolids (22) by both reducing the number of organisms available to create unpleasant or malodorous conditions and making organic substrates less suitable for bacterial metabolism and growth. Some of the mechanisms

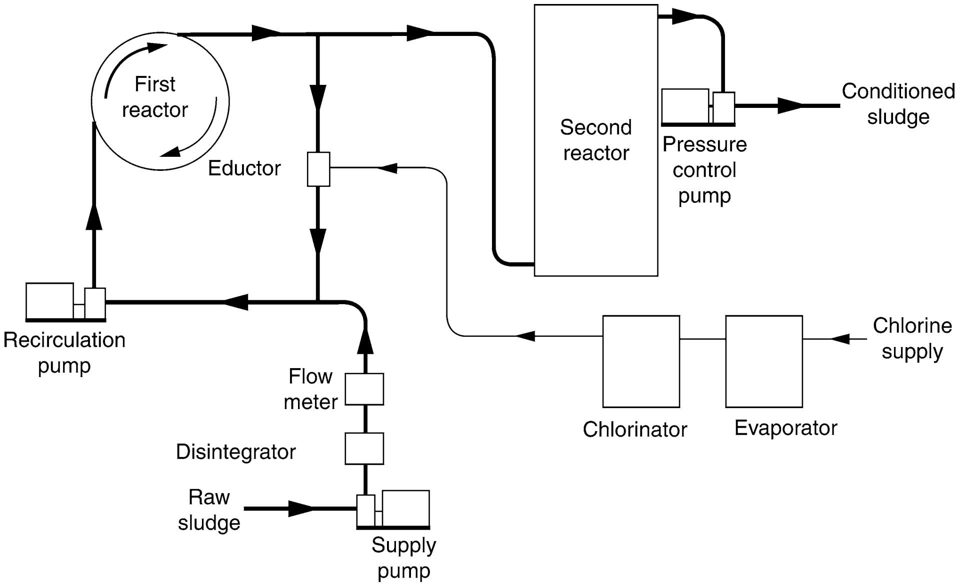

Fig. 16. Schematic diagram of a chlorine oxidation system (*Source:* US EPA).

responsible are oxidation, addition of chlorine to unsaturated compounds, and displacement of hydrogen by chlorine.

The immediate reaction from addition of gaseous chlorine to water is shown below:

$$Cl_2 + H_2O \rightarrow HOCl + H^+ + Cl^- \qquad (5)$$

In the chlorine stabilization process, sufficient acid is produced to reduce the pH of the biosolids to a range of 2–3. Dissociation of HOCl to H^+ and OCl^- is suppressed by low pH and therefore is not significant. Cl_2 and HOCl are highly reactive and powerful bactericides and viricides. The chloride ion has no disinfection capability.

The process stream immediately following the chlorine addition is substantially a chlorine solution containing biosolids. The solution contains (in molecular form) as much as 10% of the total chlorine species present. The predominant species in solution is undissociated HOCl. HOCl and Cl_2 react with biosolids to oxidize ammonia to chloramines and organic nitrogen to organic chloramines. Other reduced ions, such as Fe^{2+} and S^{2-}, are oxidized at the same time. Some of the oxidized end products, such as chloramines and organic chloramines, are germicidal and viricidal (64,65).

The chlorine stabilization unit consists of a disintegrator, a recirculation pump, two reaction tanks, a chlorine eductor, and a pressure control pump (5). A chlorine evaporator and/or a chlorinator, feed pump, and inlet flow meter can be purchased with the unit or separately. The unit is often supplied by the manufacturer as a complete package mounted on a skid plate and ready for installation. A detailed diagram of the unit is shown in Fig. 16.

In the first operating step, biosolids are pumped through a disintegrator which reduces particle size and therefore, provides more biosolids surface area for contact with the chlorine. Chlorinated biosolids from the first reactor are mixed with raw biosolids just before reaching the recirculation pump. The combined flow then passes through the first reaction tank. Chlorine is added via an eductor located in the recirculating loop. Recirculation aids mixing and efficient chlorine use. The ratio of recirculated reacted product to raw biosolids at design capacity is about 7:1. System pressure is maintained in the 30–35 psi (210–240 kN/cm^2) range, by a pressure control pump located at the discharge of the second reactor. The pressure provides a driving force to ensure penetration of chlorine into the biosolids particles. The second reactor tank increases system detention time, allowing more complete reaction between the biosolids and the chlorine.

Flow patterns within the two reactor tanks are high, in the form of velocity spirals, with tangential discharges. The tanks are oriented with the spiral axis of the first in a horizontal plane and the second in a vertical plane. Solids that settle during period of nonoperation are easily resuspended when the process is started again. The system is neither drained nor cleaned between operating period. A holding tank should be provided for feed storage and for flow equalization. Blending done in the tank also helps to maintain feed uniformity, thus providing biosolids of uniform chlorine demand and minimizing the need to frequently adjust chlorine dose. Biosolids blending is particularly valuable for processing of primary biosolids, which tend to be more concentrated when initially pumped from the sedimentation tank than at the end of the pumping cycle. Similarly, where primary and secondary biosolids are treated together, blending can be accomplished in the holding tank. However, continuously WAS might be adequately treated without prior blending, provided that solids concentration is nearly constant with time. Mixing is usually done by mechanical or air agitation. Air mixing is preferable, because it enhances aerobic conditions, reduces odor, and averts problems with fouling of the impellers by rags and strings. Odor can be controlled in the holding tank if a portion of the filtrate or supernatant from the dewatering process is returned to it.

If the chlorine demand of the liquid fraction of the biosolids is high, separation of some of the liquid from solids by thickening prior to chlorination may substantially reduce total biosolids chlorine demand. However, if the chlorine demand is low, thickening will not be beneficial. Solids concentrations above certain defined limits should not be exceeded, because the diffusion rate of chlorine through the biosolids is hindered and processing rates must be reduced to provide additional time for the chlorine to reach reaction sites. Normally, processing rates are not affected if solids concentrations are below the following values:

a. Primary biosolids or primary plus trickling filter humus: 4%.
b. Primary biosolids + WAS: 4%.
c. WAS: 1.5%
d. Processing rates for higher concentrations must be determined on a case-by-case basis.

Use of a holding tank downstream of the chlorine oxidation process allows subsequent processes to run independently and at their own best rate. Solids settling may occur in the tank after an initial period of flotation. The tank can, therefore, be used to separate the solid and liquid fractions of the stabilized product.

10.2. Advantages and Disadvantages

Chlorine oxidation has been used to treat raw and digested primary biosolids, raw and digested secondary biosolids, septage, digester supernatants, and sidestreams from dewatering processes. The chlorine stabilization process has several attractive features. It can be operated intermittently, so long as sufficient storage volume is available prior to and following the unit. Unlike biological biosolids processing systems, the process can be started up, run for a few hours, and turned off. A constant supply of process feed is not required. As a result, operating costs are directly dependent upon production rates, and costs attributable to overcapacity are eliminated.

Chlorine oxidation is a chemical process and is thus operationally insensitive to factors such as toxic materials in the biosolids, which adversely affect biological stabilization systems. It can also process feed streams of widely varying character, such as digested biosolids and digester supernatant, within a short period of time. This flexibility is not characteristic of anaerobic or aerobic digestion processes.

Disadvantages of the chlorine stabilization process center on chemical, operational, and environmental factors. From a chemical standpoint, the low pH of chlorine-stabilized biosolids may require the biosolids to be partially neutralized prior to mechanical dewatering or before being applied to acid soils. Costs of neutralization are in addition to chlorine costs. As mentioned earlier, chlorine stabilization does not reduce biosolids mass nor produce methane gas as a by-product for energy generation. The process consumes relatively large amount of chlorine.

There is concern that chlorine oxidation of biosolids, septage, and sidestreams from biosolids treatment processes would result in increased levels of toxic chlorinated organics in the treated materials (66). Measures should be taken to mitigate environmental concerns when the chlorine oxidation processes are used. These are as follows:

1. Provisions should be made to deal with the filtrate, centrate, or decant from the process, including return to the wastewater treatment plant, unless this practice leads to wastewater treatment plant upset or to violations of effluent standards; or to treat by activated carbon absorption or other means.
2. If the treated biosolids leaving the pressurized chlorinator are discharged to a tank sparged with air, the gases from the tank should be vented away from workers.
3. Treated solids should be disposed of with care. Consideration should be given to:
    a. Using secured landfills or landfills located at hydrogeographically isolated sites.
    b. Treating leachates from secured landfills to prevent contamination of surface or groundwater.
    c. Directly incorporating the solids into soil at rates sufficiently low to minimize leachate production. Direct incorporation as opposed to surface spreading should be used to prevent consumption of solids by grazing animals.
    d. Using erosion control measures to prevent runoff contaminated with toxic chlorinated compounds from entering surface water.
    e. Providing adequate monitoring of facilities to assure detection of unexpected problems.

10.3. Chlorine Requirements

Chlorine demand varies with the characteristics of each waste stream. Demand can be estimated from Table 5 for cases in which a combination of biosolids and/or sidestreams makes up the process feed. The demand of the biosolids produced by combining two streams is the weighted average of the demands of the individual streams. For example, using Table 5 one estimates that the demand of biosolids consists of five volumes of 0.7% WAS and one volume of 4% primary biosolids is about $(17 + 5 \times 7)/6 =$ 9 lb/1000 gal (1 kg/1000 L).

Table 5
Chlorine Requirements for Biosolids and Sidestream Processing

Feed stream	Suspended solids (%)	Chlorine requirement (lb/1000 gal)
Primary sludge	4	17
WAS		
With prior primary treatment	0.7	7
No primary treatment	0.7	7
From contact stabilization	0.7	7
Sludge from low and high rate trickling filters	1	10
Digester supernatant	0.3	2–10
Septage	1.2	6

Source: US EPA.

If chlorine residual is desired to provide added protection against septicity, then additional chlorine should be added in an amount equal to the required residual. For instance, if a residual of 200 mg/L chlorine is required for the biosolids combination just discussed, then an additional (200 mg/L) (8.34)/1000 = 1.7 lb of chlorine/1000 gal of biosolids (0.20 kg/1000 L) should be added, bringing the total chlorine addition to 10–11 lb/1000 gal (1.2–1.3 kg/1000 L) of biosolids. For solids concentrations other than those shown, the chlorine demand per gallon varies in proportion to the solids concentration (67). For example, if the solids concentration were to double, chlorine demand would also double.

10.4. Characteristics of Chlorine-Stabilized Materials

10.4.1. Stabilized Biosolids

Characteristics of freshly treated biosolids are a pH of 2.0–3.0 and a chlorine residual of approx 200 mg/L. Retention in a downstream holding tank allows the chlorine residual to drop to zero and the pH to rise to between 4.5 and 6.5. Normally, a slight medicinal odor is present. After adequate addition of chlorine, the color of the biosolids changes from black to light brown.

Generally, chlorine oxidation improves the sand bed dewaterability of many biosolids and septages. If properly chlorinated, the biosolids are stable and do not undergo anaerobic activity for at least 20 d. When properly disposed in landfills or on the soil, the chlorinated biosolids do not exhibit septicity during handling and disposal. If stored in lagoons, the biosolids must be sufficiently aerated to avoid odor and septicity problems, especially in warm weather.

Production of chlorinated hydrocarbons by the chlorine stabilization process has been the subject of research efforts since the process was conceived. Early studies (1971) by Metcalf and Eddy, for the BIF Company by then-current technology were aimed at the detection of specific objectionable compounds. This work indicated that, rather than producing the compounds, chlorine stabilization actually seemed to lower their concentrations in most instances. Later work using more advanced gas chromatograph-mass spectrometry techniques has revealed the production of 0.9–1% by weight organic chlorine in several biosolids stabilized by the chlorine oxidation process (66). These results indicated that as much as 10–20% of solids had chemically reacted with the chlorine. Additional studies by ASTRE for the BIF Company suggested that total identifiable chlorinated organic compounds nearly doubled when the particular raw biosolids studied were treated by a chlorine oxidation process. A six-fold increase was found in the amount of chlorinated organic toxic compounds following chlorine oxidation and an eightfold increase in the amount of total organic toxic compounds.

10.4.2. Supernatant/Filtrate/Subnatant Quality

These process streams are produced by thickening and/or dewatering operations after chlorine treatment. Filtrates from sand bed dewatering are typically clear and colorless. The pH varies from 4 to 6, and no residual chlorine remains. Generally, filtrate from chlorine-treated biosolids contains lower suspended solids and BOD_5 than the filtrate produced when filtering digested biosolids. Typical filtrate composition is 50–150 mg/L suspended solids and 100–300 mg/L BOD_5 with low turbidity and color.

In bench-scale studies simulating the chlorine oxidation process, Oliver et al. (68) found that acidic conditions enhanced the release of heavy metals from biosolids. Sukenik et al. (69) noted an increase in supernatant chemical oxygen demand (COD) after biosolids treatment by chlorine oxidation. Though the reason for the increase is undetermined, the suggestion was made that chlorine may solubilize the oxygen-demanding material rather than oxidize it. Generally, biochemical oxygen demand of the supernatant is comparable with that of raw wastewater. Data collected at Alma, MI, indicate that chemically precipitated phosphate is not redissolved by the chlorine oxidation process (67). A US EPA report (66) indicated that chlorinated organics were present in the centrate from several chlorinated biosolids samples Although less than 0.5% of the organic compounds assumed to be present could be identified, eight chlorinated compounds on the list of toxic substances were detected, including three known or suspected carcinogens.

10.5. Costs

Capital costs for chlorine stabilization systems tend to be less than for conventional anaerobic digestion systems of equal capacity. Normally, the system is furnished by the manufacturer on a skid-plate and in a ready-to-install condition. Table 6 shows actual 1979 capital costs for systems of specified capacity for two different feed biosolids:

a. Capital costs were derived by Purifax from actual installations in 1979 USD (Cost Index for utilities = 257.20, Appendix). To determine current costs in 2006 USD (Cost Index for utilities = 528.12) multiply capital costs by a factor of 528.12/257.20 = 2.05 (24).
b. Costs include chemical oxidizer chlorine, biosolids macerator, biosolids feed pump, motor starters, vacuum-type chlorinator, freight, and start-up service.

Because the chlorine stabilization process can be operated intermittently, annual operating costs are proportional to the quantity of material processed. Table 7 displays the operating cost data. Chlorine unit costs vary with annual usage, method of transportation and transportation distances, and competition. The reader should note the following:

a. Cost data in Table 7 were derived by Purifax from actual installations in 1974 USD (Cost Index for utilities = 170.45, Appendix). To determine current costs in 2006 USD (Cost Index for utilities = 528.12) multiply costs by a factor of 528.12/170.45 = 3.10 (24).

Table 6
Chlorine Stabilization Capital Costs (1979)

Capacity (gal/h)		
Primary and WAS[a]	WAS only[b]	Budgetary cost (USD)
660	960	82,000
1320	1800	137,000
2940	4200	175,000
5880	8520	228,000
13,080	18,300	307,000

Source: US EPA.

[a] Solids concentration 3% by weight.

[b] Solids concentration 1.5% by weight.

Table 7
Actual Operating Costs for Chlorine Stabilization System

	Chlorine		Cost (USD/T of dry solids)		
Process stream and year reported	Dosage (lb/t of dry solids)	Cost (cents/lb)	Chlorine	Power	Chlorine and power
Primary and WAS					
Ravena-Coeymans, NY (1974)	167	11.35	18.95	1.90	20.85
Plainfield, CT, (1973)	148	14	20.72	2.07	22.79
Extended aeration					
Plainfield, CT (1975)	180	14	25.20	2.52	27.72
WAS only					
Fair lawn, NJ	211	9.85	20.78	2.08	22.86

Source: US EPA.

b. Although it is not related to the cost of chlorine stabilization of biosolids, additional chemical costs can result if chemical conditioning is necessary before mechanical dewatering. Chemical conditioning of chlorine-stabilized biosolids consists of adding sodium hydroxide or lime to raise the pH to between 4.5 and 5.5 and then adding the proper dosage of an appropriate coagulant. Although more expensive, sodium hydroxide is generally preferred to lime because it reacts faster. Neutralization can be done in-line, without need of an intermediate detention tank.

c. Sodium hydroxide requirements range from 20 to 30 lb/t of dry solids (10–15 kg/T) for primary biosolids to 10–20 lb/t of dry solids (5–10 kg/t) for secondary biosolids. At 2004 cost of 21 cents/lb (46 cents/kg) this is equivalent to a cost 2.1–6.2 USD/t (2.3–6.8 USD/T) of dry solids. Polymer costs are equivalent to those required for dewatering of biosolids stabilized by other means and are generally more than the cost of pH adjustment.

d. 2006 Costs for neutralizing chlorine-stabilized biosolids before to their spreading on acid soils are about 1.56–2.33 USD/t (1.63–2.57 USD/T), assuming that 20–30 lb of $Ca(OH)_2$ are required per ton (10–15 kg/t) of stabilized solids and $Ca(OH)_2$ costs are 0.08 USD/lb (0.19 USD/kg).

e. Power costs of operating the stabilization system are estimated at 10% of chlorine costs. Additional power costs are incurred if mixing is used in the holding tank upstream from the stabilization process.

f. Labor costs are incurred only for daily start-up, shutdown, periodic checks, and maintenance, and are small in comparison to other operating costs.

11. DESIGN EXAMPLE

A designer has calculated that the rotary drum, cloth belt vacuum filter that will be used at the plant must be capable of dewatering a maximum of 272 kg/h (600 lb/h) of biosolids. The biosolids stream will be a mixture of 40% primary and 60% WAS, and it will be anaerobically digested. The vacuum filter is to operate 7 h/d, 5 d/wk. To design for a margin of safety in the chemical feed equipment, the designer has used the higher values shown in Table 3. Chemical feeders should be capable of adding 60 kg/T (120 lb/t) of $FeCl_3$ and 210 kg/T (420 lb/t) of CaO.

Solution

Maximum daily amount of biosolids to be dewatered is:

272 kg biosolids/h × (7 h/d) = 1904 kg/d (4200 lb/d).

Maximum amount of $FeCl_3$ required per day is:

1904 kg biosolids/d × 60 kg $FeCl_3$/1000 kg biosolids = 114 kg/d (252 lb/d).

The $FeCl_3$ is available as a 40% solution that contains 1 kg of active ingredient per 1.77 L of solution (4.72 lb/gal of solution).

114 kg/d × 1.77 L of product/1 kg $FeCl_3$ = 202 L of $FeCl_3$
solution needed/d (53.4 gal/d).

Maximum amount of CaO required per day is:

1904 kg biosolids/d × 210 kg CaO/1000 kg biosolids = 400 kg CaO/d (882 lb/d).

The quicklime is available at 90% CaO:

400 kg CaO/d × 1 kg quicklime/0.9 kg CaO = 445 kg quicklime/d (980 lb/d).

The amount of extra biosolids produced because of chemical addition is estimated at 1 kg for every kg of $FeCl_3$ and quicklime added. Therefore, total maximum daily dry solids to be disposed of are:

1905 kg biosolids + 114 kg $FeCl_3$ + 445 kg quicklime = 2464 kg (5432 lb) of solids.

This is the equivalent of 12,320 kg (27,160 lb) of wet solids (biosolids) at a minimum of 20% solids.

Cost associated with this amount of chemicals in 2006 USD:

$FeCl_3$ = 0.39 USD/kg (0.18 USD/lb).
Quicklime = 0.10 USD/kg (0.046 USD/lb).
114 kg $FeCl_3$/d × 0.39 USD/kg = 44.46 USD/d.
445 kg quicklime/d × 0.10 USD/kg = 44.50 USD/d.
1905 kg biosolids/d/t/1000 kg = 1.9 T/d (2.1 t/d).
[(44.46 USD + 44.50 USD)/d]/(1.9 T/d) = 47 USD/dry T (43 USD/t).

NOMENCLATURE

A	Area of filter (m^2)
b	Slope of time/volume vs volume curve (s/cm^6)
BOD_5	5-d biochemical oxygen demand produced by VSS solubilization (kg or lb)
P	Pressure of filtration (N/m^2)
PB	Primary biosolids (dry kg or lb)
r	Specific resistance (m/kg)
t	ton (English unit)
T	tonne (Metric unit)
VSS	Volatile suspended solids solubilized (dry kg or lb)
W	Weight of dry solids/volume of filtrate (kg/m^3)
WAS	Waste-activated sludge (Secondary biosolids) (dry kg or lb)
μ	Viscosity of filtrate (N s/m^2)

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APPENDIX

United States Yearly Average Cost Index for Utilities (24)

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

12

Elutriation and Polymer Conditioning

Lawrence K. Wang, Shoou-Yuh Chang, Yung-Tse Hung, and J. Paul Chen

CONTENTS

1. ELUTRIATION PROCESS DESCRIPTION

Elutriation is described by Wang (1) as “A process of sludge conditioning whereby a sludge is washed either by fresh water or plant effluent, to reduce the sludge alkalinity and fine particles, thus decreasing the amount of required coagulant in further treatment steps, or in sludge dewatering,” as shown in Fig. 1. Elutriation is the term commonly used to refer to the washing of anaerobically digested sludge before vacuum filtration, as shown in Fig. 2. Washing causes a dilution of the bicarbonate alkalinity in the sludge and therefore reduces the demand for acidic metal salt by as much as 50% (2–14).

The process itself was patented by Genter in 1941 (3,4). Although it typically uses one or two tanks, any number of tanks can be used. Two to six volumes of washwater, typically plant effluent, flow countercurrent to one volume of anaerobically digested sludge. Elutriation tanks are designed to act as gravity thickeners, with a mass solids loading of 8–10 lb/ft^2/d (39–48.8 kg/m^2/d).

During the process of anaerobic digestion of sewage sludge much of the organic carbon content is converted to methane (CH_4) and to carbon dioxide (CO_2) gases. These gases are usually collected in the space over the digester content surface. Therefore, there exists a substantial concentration of the gases in the digester contents. The actual concentration can be estimated as shown later. The gas of interest in this discussion is CO_2, which exists in solution in equilibrium with the bicarbonate ion (HCO_3^-), which in turn

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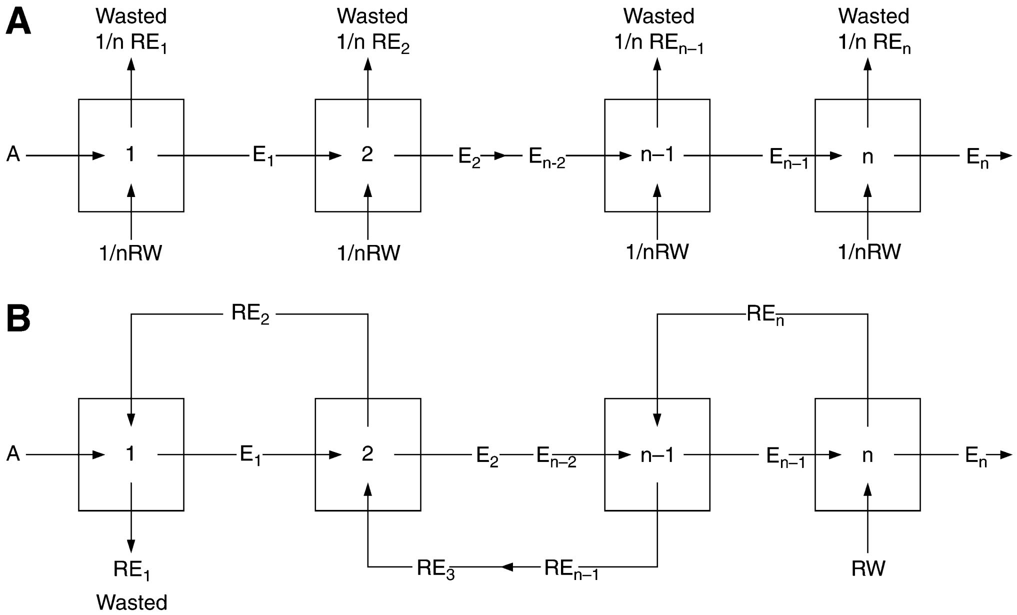

Fig. 1. Sludge washing or elutriation. **(A)** Top: multiple elutriation in a single tank (**B**) Bottom: Countercurrent elutriation in multiple tanks (4).

is in equilibrium with the carbonate ion (CO_3^{2-}). The distribution among CO_2–HCO_3^-–CO_3^{2-} is dependent somewhat upon temperature and total dissolved solids but primarily upon pH. In practical usage the sum of the bicarbonate ion plus the carbonate ion is termed alkalinity. For well-digested sludges the pH is about 7.0 (6.8–7.2) and at this pH almost all of the alkalinity is in the form of the bicarbonate ion. Carbon dioxide is about 30% of the total gas produced.

According to Fair et al. (4) bicarbonate ion concentration in digested sludge may be as much as 100 times the concentration in the fresh sludge;which is approximately the quantity in the water supply. Because even digested sludge does not easily dewater, particularly in the brief exposure time in a vacuum filter or centrifuge, it is usually necessary to add coagulating chemicals to aid the dewatering process (15–27).

2. ELUTRIATION PROCESS DESIGN CONSIDERATIONS

2.1. Reactor Design Considerations

The losing solids problem of elutriation process can be solved, and 90–92% capture achieved, with the use of polymers. Recommended current elutriation design considerations are listed here:

a. Tanks should be loaded at hydraulic loadings (total of both sludge and washwater flow) of 200–300 gal/d/ft^2 (69–104 L/d/m^2) and solids loading of 8–15 lb/d/ft^2.
b. Tanks should have the best possible inlet structure to minimize inlet momentum.
c. Baffling should be used to prevent tank currents.
d. Tanks should be provided with scum collection.
e. Polymer addition should be provided.

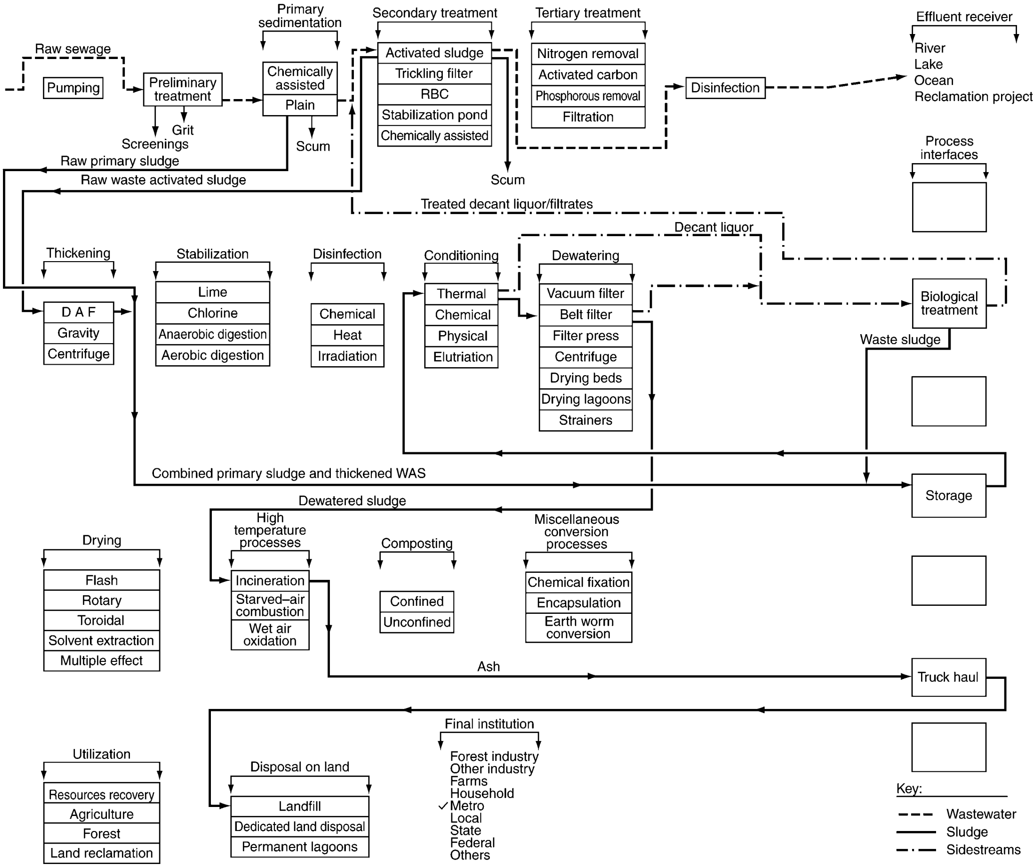

Fig. 2. Flowsheet for process selection in wastewater and sludge treatment (*Source:* US EPA).

2.2. *Elutriate Disposal Considerations*

In addition to reducing alkalinity, the elutriation process will also wash out 10–45% of the solids from the incoming sludge stream. Elutriate was recycled back to the main plant and eventually degraded in the plant effluent (6–11). Other elutriation applications and elutriate disposal technologies are also reported by the researchers (29,30).

2.3. *New Technology Considerations*

There are two conventional elutriation process systems (2,11) for continuous sludge washing:

a. Multiple elutriation in a single tank.
b. Countercurrent elutriation in multiple tanks.

Section 3 presents the procedures for design of the earlier two conventional elutriation processes. The readers are referred to other sources of information on recent development in elutriation technology (11–14). Beckman Coulter, Inc. (12) has developed a “batch elutriation” process equipment which is commercially available. 5R Research Inc. (13) has developed a magnetic elutriation technology for clean and efficient processing of iron ore. Whether or not the magnetic elutriation technology can be adopted for activated sludge elutriation remains unknown. Wang (15,16) has applied similar magnetic technology for sludge thickening with successful results. Zimmels (14) describes the performance criteria of a modern elutriation process, and also introduces the following new elutriators: (a) Centrifugal elutriators. (b) Counterflow gravity elutriators. (c) Axial crossflow elutriators and (d) Electromagnetic elutriators.

Hwang et al (29) report an Anaerobic Digestion Elutriated Phase Treatment (ADEPT) for resource recovery and nitrogen removal from slurry-type piggery waste (29). Ahn and Speece (30) report a process comprising an upflow sludge blanket-type high performance elutriated acid fermenter for organic acids recovery, followed by an upflow-type crystallization reactor using waste lime for nutrient recovery.

2.4. Benefit

The benefits of elutriation process are many. Renoux and coworkers used the Microtox® assay (*Vibrio fischeri*) to determine if the handling of environmental or industrial samples affects their toxicity. The Microtox® system is a rapid screening tool for a variety of toxicity testing, developed by AZUR Environmental and Strategic Diagnostics Inc., Newark, Delaware, USA (28). It was found that irradiation and 3 mo freezing of sewage sludge elutriates decreased the toxicity of these samples as measured using the Microtox test. The effect of storage at –20°C was also confirmed with the lettuce Lactuca sativa root elongation assay (17).

3. ELUTRIATION PROCESS DESIGN PROCEDURES

To eliminate, or at least to, reduce the bicarbonate ion demand, mixing of the sludge with water will result in a distribution of the bicarbonate between the washwater and the sludge. This is the purpose of the process of elutriation (11). The bicarbonate ion may also be reduced by precipitation similar to calcium carbonate ($CaCO_3$), by using lime. This will result in large amounts of precipitate. Often a combination of elutriation and lime precipitation is used in conjunction with the trivalent metallic coagulants.

Genter (4) has suggested for ferric chloride ($FeCl_3$) that

$$p_c = [1.08 \times 10^{-4} AP/(100 - p)] + 1.6 p_v/p_f \tag{1}$$

where A is the alkalinity of the sludge moisture in milligram per liter as $CaCO_3$, and p_c, p, p_v, and p_f are respectively the percentages of chemical ($FeCl_3$), moisture, volatile matter and fixed solids in sludge all on a dry basis. The ratio of p_v/p_f is significantly reduced by digestion. However, the value of A is greatly increased by anaerobic sludge digestion. Sludge elutriation can be a single step or multiple steps conducted in single or multiple tanks. During washing the water-sludge mixture is kept in suspension by mechanical means or by air mixing. The multiple step process conducted in multiple tanks is usually a countercurrent operation. Fair et al. (4) and Granstrom (11) describe the possible processes in two categories:

a. Multiple elutriation in a single tank.
b. Countercurrent elutriation in multiple tanks.

3.1. Multiple Elutriation in a Single Tank

In the process shown in Fig. 1A, elutriating water is added to the sludge in a tank and is mixed well. The sludge is allowed to settle and the supernatant is drawn off. The removed supernatant is then replaced with fresh water and the process repeated until the desired level of alkalinity in the surface is reached (11). Let, R, ratio of the total volume of elutriating water to the volume of sludge; W, alkalinity of the elutriating water, mg/L as $CaCO_3$; A, alkalinity of the sludge before elutriation, mg/L as $CaCO_3$; E, alkalinity of the elutriated sludge mg/L as $CaCO_3$; n, number of steps in the washing process. Then $1/n$ of the elutriating volume is used in each step. A mass balance of the alkalinities entering and leaving the tanks in each step can be formulated as follows:

The subscripts refer to the step, 1–n

$$\begin{aligned}
&(1/n)RW + A = (1/n)RE_1 + E_1\\
&(1/n)RW + E_1 = (1/n)RE_2 + E_2\\
&\qquad\vdots\\
&(1/n)RW + E_{n-2} = (1/n)RE_{n-1} + E_{n-1}\\
&(1/n)RW + E_{n-1} = (1/n)RE_n + E_n
\end{aligned}$$

The value of E_n can thus be calculated in a stepwise operation or by use of Eq. (2);

$$E_n = \frac{A + W\left(\frac{R}{n} + 1 - 1\right)^n}{\left(\frac{R}{n} + 1\right)^n} \tag{2}$$

or

$$R = n\left(\frac{A - W}{E_n - W}\right)^{1/n} - 1 \tag{3}$$

3.2. Countercurrent Elutriation in Multiple Tanks

In the countercurrent elutriation, shown in Fig. 1B, the sludge and elutriating water move through a series of n tanks in countercurrent manner. This has the distinct advantage of requiring less elutriating water for the same reduction in sludge alkalinity. In this scheme, the clean wash water is introduced into the last tank, i.e., with the partially elutriated sludge. The relatively higher in alkalinity washwater meets the less elutriated sludge. Similarly, as earlier, a mass balance of alkalinities can be formulated for n tanks (11).

$$\begin{aligned}
&RW + E_{n-1} = RE_n + E_n\\
&RE_n = E_{n-2} = RE_{n-1} + E_{n-1}\\
&RE_3 + E_1 = RE_2 + E_2\\
&RE_2 + A = RE_1 + E_1
\end{aligned}$$

The value of E_n can be computed in a stepwise fashion or by use of Eq. (4):

$$\begin{aligned}
E_n &= \frac{A + W\left(R^n + R^{n-1} \ldots\ldots + R\right)}{R^n + R^{n-1} \ldots\ldots + R + 1}\\
&= \frac{\mathrm{A}\left(\mathrm{R}-1\right) + WR(R^n - 1)}{R^{n+1} - 1}
\end{aligned} \tag{4}$$

or

$$(R^{n+1} - 1)/(R - 1) = (A - W)/(E_n - W) \tag{5}$$

4. CHEMICAL CONDITIONING WITH SOLUBLE ORGANIC AND INORGANIC POLYMERS

During the past decade, important advances have been made in the manufacture of soluble polymers for use in wastewater sludge treatment. Polymers are now widely used in sludge conditioning. These materials differ greatly in chemical composition, functional effectiveness, and cost-effectiveness (11). Zimmels (14), introduces application of polymers in elutriation process. Selection of the correct soluble polymer requires that the designer work with soluble polymer suppliers, equipment suppliers, and plant operating personnel. Evaluations should be made on site and with the sludges to be conditioned. Since new types and grades of polymers are continually being introduced, the evaluation process is an ongoing one.

Soluble organic polymers can be divided into three categories:

a. Nonionic polymers.
b. Anionic polymer (anionic polyelectrolytes).
c. Cationic polymer (cationic polyelectrolytes).

Technically speaking, only cationic and anionic polymers can be called polyelectrolytes because they bear charges. However, in the engineering field most of the technicians do not know the difference between polymers and polyelectrolytes, so the terms of “polymers” and “polyelectrolytes” are used interchangeably.

4.1. Soluble Nonionic Organic Polymers

Organic soluble polymers are long chained water soluble, specialty chemicals. They can either be completely synthesized from individual monomers, or they can be made by the chemical addition of functional monomers, or groups, to naturally occurring polymers. A monomer is the subunit from which polymers are made through various types of polymerization reactions. The backbone monomer most widely used in synthetic organic polymers is acrylamide. Polyacrylamide, created when the monomers combine to form a long, thread-like molecule with a molecular weight in the millions, is shown on Fig. 3. The polyacrylamide shown in the form is essentially nonionic. That is to say it carries no net electrical charge in aqueous solutions. However, under certain conditions and with some solids, the polyacrylamide can be sufficiently surface-active to perform as a flocculant.

4.2. Soluble Ionic Organic Polymers (Polyelectrolytes)

Anionic-type polyacrylamide flocculants carry a negative electrical charge in aqueous solutions and are made by either hydrolyzing the amide group (NH_2) or combining the acrylamide polymer with an anionic monomer.

Cationic polyacrylamides carry a positive electrical charge in aqueous solutions and can be prepared by chemical modification of essentially nonionic polyacrylamide or by combining the cationic monomer with acrylamide. When cationic monomers are copolymerized with acrylamide in varying proportions, a family of cationic polyelectrolytes with varying degrees of charge is produced. These polyelectrolytes are the most widely used polymers for sludge conditioning, because most sludge solids carry a negative charge. The characteristics of the sludge to be processed and the type of thickening

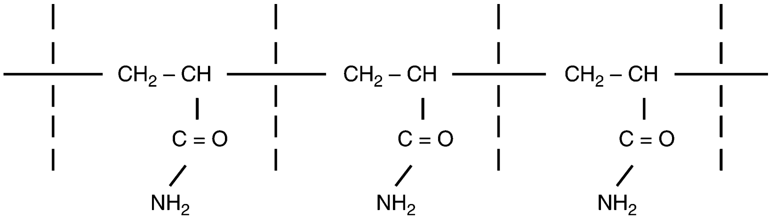

Fig. 3. Polyacrylamide molecule—backbone of the synthetic organic polyelectrolytes (*Source:* US EPA).

Table 1
Representative Dry Powder Cationic Polyelectrolytes

Type	Relative cationic density	Molecular weight	Dosage (lb/t dry)
Polyacrylamide copolymer	Low	Very high	0.5–10
Polyacrylamide copolymer	Medium	High	2–10
Polyacrylamide copolymer	High	Medium high	2–10
Polyamine homopolymer	Complete	High	2–10

Source: US EPA. 1 t = 1 ton (English unit) = 2000 lb.

or dewatering device used will determine which of the cationic polyelectrolytes will work the best and still be cost effective. For example, an increasing degree of charge is required when sludge particles become finer, when hydration increases, and when relative surface charge increases.

Cationic polyelectrolytes are available as dry powders or liquids. The liquids come as water solutions or emulsions. The shelf life of the dry powders is usually several years, whereas most of the liquids have shelf lives of 2–6 mo and must be protected from wide ambient temperature variations in storage. Representative dry cationic polyelectrolytes are described in Table 1. They were produced for the conditioning of primary sludges or easy-to-condition mixed sludges. The incentive to produce polymers of higher positive charge resulted largely from efforts to cope with mixed sludges containing large quantities of biomass.

Relatively low-molecular weight liquid cationics with a 30–50% solids content are also available. They have been gradually displaced by the higher cationic functionality, high molecular weight, and newer, less costly liquid cationics. The various liquid cationics, in either dissolved or emulsion form, are described in representative fashion in Table 2. These liquid cationics eliminate the dustiness inherent in some dry powders but also require much more storage space. The selection of a dry, liquid, or emulsion form material usually depends on a comparison of cost-effectiveness, ease of handling, and storage requirements.

Organic polyelectrolytes dissolve in water to form solutions of varying viscosity. The resulting viscosity depends on their molecular weight and degree of ionic charge. At infinite dilution, the molecule assumes the form of an extended rod because of the repulsive effect

Table 2
Representative Liquid Cationic Polyelectrolytes

Type	Molecular weight	Solids (%)
Mannich product	Low	20
Tertiary polyamine	Low	30
Quaternary polyamine	Very low	50
Cationic homopolymer	Low to medium	16–20
Emulsion copolymer	Low to medium	25–35

Source: US EPA.

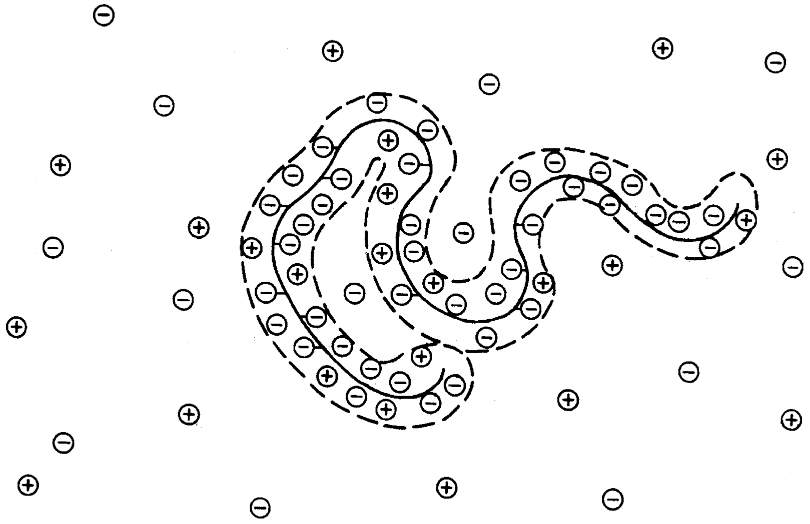

Fig. 4. Typical configuration of a cationic polyelectrolyte in solution (*Source:* US EPA).

of the adjacent-charged sites along the length of the polymer chain. At normal concentrations the long thread-like charged cationic polyelectrolyte has the shape of a random coil, as shown on Figs. 4 and 5. However, this simplified drawing neither shows the tremendous length of the polymeric molecular chain nor does it illustrate the very large number of active polymer chains that are available in a polymer solution. It has been estimated that a dosage of 0.2 mg/L of polyelectrolyte having a molecular weight of 100,000 would provide 120 trillion active chains/L of water treated.

4.3. Polyelectrolyte Conditioning Process for Sludge Thickening

Thickening and dewatering are inhibited by the sludge particles, chemical characteristics, and physical configurations. Polyelectrolytes in solution act by adhering to the sludge particle surfaces thus causing:

a. Desorption of bound surface water.
b. Agglomerization of small particulates by bridging between particles (18).

The result is the formation of a permeable sludge cake matrix which is able to release water. Figure 5 illustrates the polyelectrolyte-solid attachment mechanism. The first two

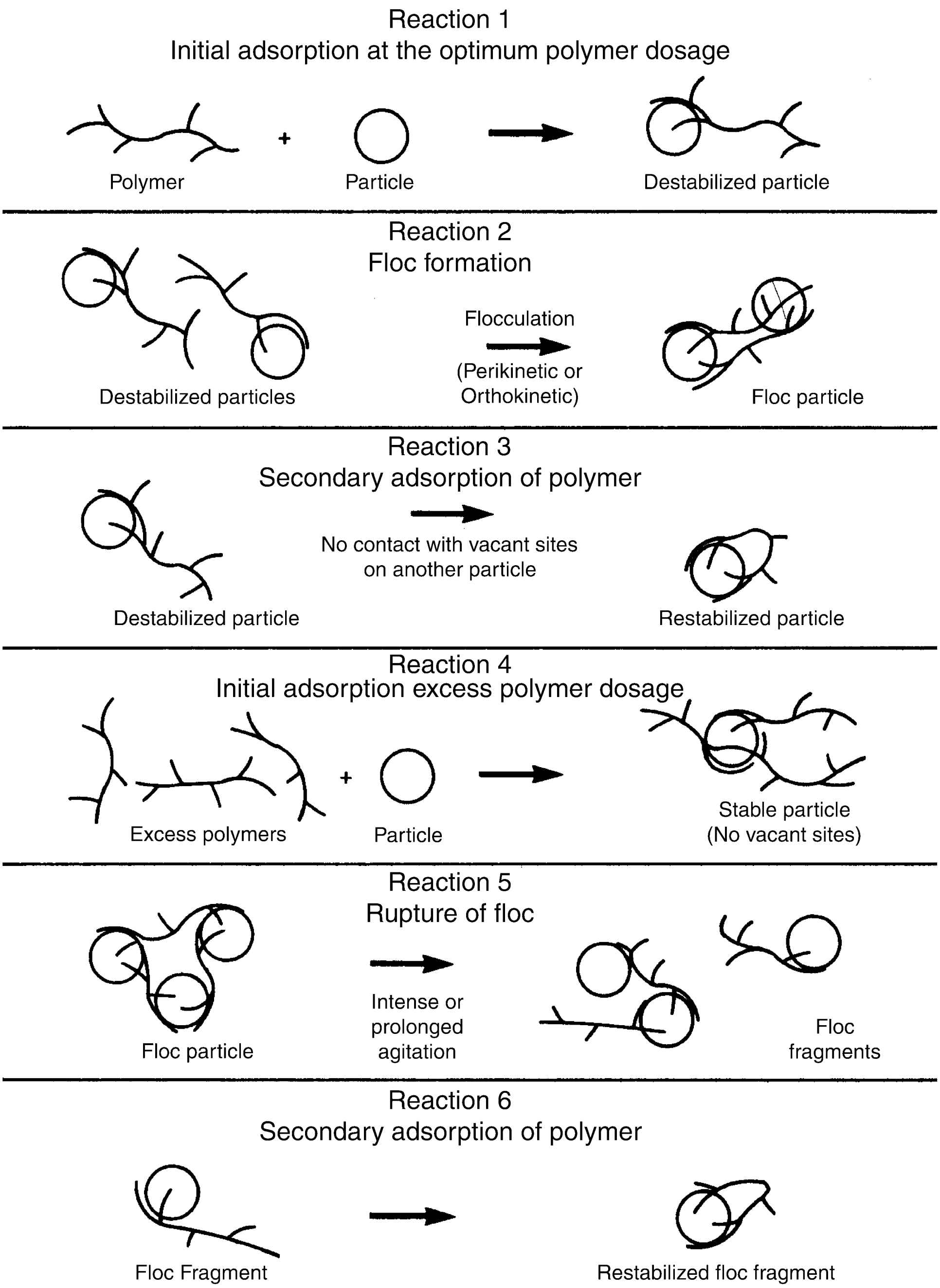

Fig. 5. Schematic representation of the bridging model for the destabilization of colloids by polymers (*Source:* US EPA).

reactions noted on Fig. 5 are the desirable ones and represent what occurs in normal practice. The other four reactions represent what can occur from overdosage or too much shear of flocculated sludge. The problems reflected in reactions three through six rarely occur with a well-designed process. Normally, the addition of polyelectrolyte is not considered in the original design because of operating cost, but it has been used to upgrade existing facilities. Experience to date has indicated that the addition of polyelectrolyte to a thickener will:

a. Give higher solids capture than a unit not receiving polymer addition.
b. Allow solids loading rate two to four times more than a unit not receiving polymer addition.
c. Maintain the same underflow solids concentration as a unit not receiving polymer addition.

When polyelectrolyte is used to condition sludge for gravity thickening, it should be added into the sludge feed line. The point of addition should provide good mixing and not cause excessive shear before the conditioned sludge discharges into the sludge feed well. When polyelectrolyte conditioning is used with centrifugal thickening of sludge, several points of addition should be provided. The optimum point of addition is influenced by differences in polymer charge densities, required polymer sludge reaction times, and sludge characteristics.

Recommended points of addition are as follows:

a. Directly before the inlet side of the sludge feed pump.
b. Immediately downstream of the sludge feed pump.
c. To the centrifuges' sludge feed line and just before its connection to the centrifuge.

4.4. Polyelectrolyte Conditioning Process for Dewatering

Various dewatering methods are discussed elsewhere in the book. Polyelectrolytes were originally used to condition primary sludges and easy-to-dewater mixtures of primary and secondary sludges, for dewatering by rotary vacuum filters or solid bowl decanter centrifuges. Improvement in the effectiveness of polyelectrolytes has led to their increasing use with all types of dewatering processes. Reasons for selecting polyelectrolytes over inorganic chemical conditioners are as follows:

a. Little additional sludge mass is produced. Inorganic chemical conditioners typically increase sludge mass by 15–30%.
b. If dewatered sludge is to be used as a fuel for incineration, polyelectrolytes do not lower the fuel value.
c. They allow for cleaner material-handling operations.
d. They reduce operation and maintenance problems (18).

The majority of municipal vacuum filtration processes in the United States still dewater sludge conditioned with ferric chloride and lime. However, several facilities have begun using polyelectrolytes for conditioning and have realized cost is reduced owing to less equipment maintenance, fewer materials handling problems, and reduction of cost in downstream sludge processing operations. Table 3 shows additional levels of dry polyelectrolyte used in conditioning different types of sludge for vacuum filtration. When using polyelectrolyte conditioning prior to vacuum filtration, the designer should be aware that sludge formation properties can be quite different from those of inorganic chemical conditioners. More operator attention may be required to obtain good cake release from the cloth. Cake dryness will probably be 10–15% lower and the volatile content of the dry cake will be significantly higher than if the sludge had been conditioned with ferric chloride and lime.

Table 3
Typical Polyelectrolyte Additions for Various Sludges

Sludge type	Pounds of dry polymer added per ton of dry solids
Raw primary	0.5–1
Waste activated	8–15
Anaerobically digested primary	1.5–4
Primary plus trickling filter	2.5–5
Primary plus air waste activated	4–10
Primary plus oxygen waste activated	4–8
Anaerobically digested (primary plus air waste activated)	5–12

Source: US EPA.
Data supplied by equipment manufacturers.
1 lb/t = 0.5 kg/T.
1 t = 1 ton (English unit) = 2000 lb; 1 T = 1 tonne (metric unit) = 1000 kg

4.5. Inorganic Polymer Conditioning Process for Thickening and Dewatering

Salts of trivalent metallic cations, aluminum and iron, are also commonly used, often in conjunction with lime (CaO). Because of the high partial pressure of CO_2, in the sludge a there is high concentration of bicarbonate ion. Before the iron or aluminum salts can effectively aid in the dewatering of the sludge, a large portion of the bicarbonate ion demand for the trivalent ion must be met. This results in very large demand for the coagulating chemicals that in turn results in large amount of hydrous aluminum or ferric oxide solids (19).

4.6. Polyelectrolytes Determination and Process Control

Polyelectrolytes are used widely in chemical conditioning, sludge thickening, and sludge dewatering (31). The readers are referred to the literature for the information on polyelectrolytes determination (15,16,20–25) and process control (26,27).

5. DESIGN EXAMPLES

5.1. Example 1

A digested sludge with 50% volatile solids and 92% moisture and an alkalinity of 3500 mg/L is to be dewatered in a vacuum filter with ferric chloride pretreatment. Determine the amount of ferric chloride in pounds per thousand gallons (1 lb/1000 gal = 0.12 kg/1000 L) of sludge if there is no elutriation (11).

By use of Genter's equation

$$p_c = (1.08 \times 10^{-4}\ \mathrm{A}p/[100 - p]) + 1.6\ p_v/p_f$$
$$= 1.08 \times 10^{-4} \times 3500 \times 92/8 + 1.6 \times 50/50$$
$$= 5.95\% \text{ of } FeCl_3 \text{ on a dry weight basis, in the sludge.}$$

The weight of solids in 1000 gal of sludge with a specific gravity of 1.04 is:

weight of solids = 1000 × 8.34 × 1.04 × 0.08
= 694 lbs
= 315 kg.

The weight of $FeCl_3$ to be added to each 1000 gal (3785 L) of sludge is:

weight of $FeCl_3 = 0.0595 \times 694$

$= 41.3$ lbs

$= 18.8$ kg.

5.2. Example 2

The same digested sludge with 50% volatile solids and 92% moisture and a alkalinity of 3500 mg/L is to be dewatered on a vacuum filter (11). If the alkalinity of the sludge is reduced to 500 mg/L, determine the amount of ferric chloride/1000 gal (3785 L) of the sludge.

$p_c = (1.04 \times 10^{-4} \times 500 \times 92/8) + 1.6 \times 50/50$

$= 2.2\%$ of $FeCl_3$ on a dry weight basis, in the sludge

weight of $FeCl_3 = 0.022 \times 694$

$= 15.3$ lbs

$= 6.95$ kg.

5.3. Example 3

If the same digested sludge (50% volatile solids; 92% moisture; 3500 mg/L sludge alkalinity before elutriation; and 500 mg/L sludge alkalinity after washing twice) is to be washed twice in a batch process in a single tank with wash water having an alkalinity of 50 mg/L, what will be the ratio of the wash water to sludge? (11).

Using a single tank Eq. (3):

$$R = n\left\{\left(\frac{A-W}{E_n-W}\right)^{1/n} - 1\right\}$$

$$= 2\left\{\left(\frac{3500-50}{500-50}\right)^{1/2} - 1\right\}$$

$$= 3.54$$

5.4. Example 4

If the same digested sludge (50% volatile solids; 92% moisture; 3500 mg/L sludge alkalinity before elutriation; and 500 mg/L sludge alkalinity after countercurrent washing) is to be washed in a two tank countercurrent flow system with wash water having an alkalinity of 50 mg/L, what will be the ratio of the wash water to the sludge? (11).

Using the countercurrent flow equation Eq. (5):

$$\frac{R^{n+1}-1}{R-1} = \frac{A-W}{E_n-W}$$

$$= \frac{3500-50}{500-50}$$

$$(R^3-1)\,(R-1) = 7.67$$

or,

$$R^2 + R + 1 = 7.67$$

$$R = 2.14$$

The countercurrent system reduces the required wash water volume by,

$$\frac{3.54 - 2.14}{3.54} = 39\%$$

NOMENCLATURE

A Alkalinity of sludge before elutriation in mg/L as $CaCO_3$
E Alkalinity of elutriated sludge in mg/L as $CaCO_3$
n Number of steps in the washing process
p Percentage of moisture on a sludge dry weight basis
p_c Percentage of chemical, $FeCl_3$ on a sludge dry weight basis
p_f Percentage of fixed solids on a sludge dry weight basis
p_v Percentage of volatile matter on a sludge dry weight basis
R Ratio of the total volume of elutriating water to the volume of the sludge
W Alkalinity of the elutriating water in mg/L as $CaCO_3$

ACKNOWLEDGMENTS

This chapter, Elutriation and Polymer Conditioning, has been written in honor of Prof. Marvin L. Granstrom who was the former Chairman of the Department of Civil and Environmental Engineering, Rutgers University, Piscataway, New Jersey. Prof. Granstrom received his PhD degree from Harvard University and has educated over 100 PhD students from around the world in his life long teaching career at Rutgers. Dr. Lawrence K. Wang is among those graduates from Rutgers, who have received supervision and guidance from Dr. Granstrom. The authors of this chapter salute this great professor for the dedication of his entire academic career to promote environmental engineering education.

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13
Drying Beds

Lawrence K. Wang, Yan Li, Nazih K. Shammas,
and George P. Sakellaropoulos

CONTENTS

1. INTRODUCTION

Although numerous techniques fulfill the basic functional definition of dewatering, they do so to widely varying degrees. It is important to keep this fact in mind when comparing different dewatering and/or drying devices. For example, sludge drying beds and evaporation lagoons can be used not only to dewater a particular sludge, but also to dry it to a solids concentration of more than 50–60%. Depending on the particular device involved, dewatered sludge from a mechanical device might vary from a wet, almost flowable form, to a harder and more friable form (1–5).

Both sludge drying beds and sludge evaporation lagoons (also known as sludge drying lagoons) are the most widely used methods of municipal sludge dewatering/drying in the United States. At the present time, two-thirds of all United States wastewater treatment plants utilize drying beds and 50% of all the United States municipal sludges are dewatered by this method. Although the use of drying beds might be expected in smaller plants and in the warmer sunny regions, they are also used in several large facilities in northern climates. Precipitation and evaporation rates are the important controlling factors of the two natural drying processes (6,7). The aim of this chapter is to cover the subject of sludge drying using sludge drying beds.

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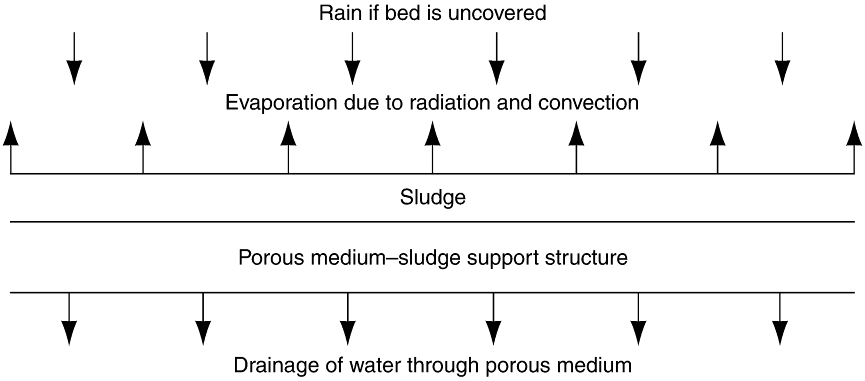


Fig. 1. Schematic of sludge dewatering in drying bed system.

2. SLUDGE DRYING BEDS PROCESS DESCRIPTION

Since the early 1900s research into the dewatering of sludge by drying beds has been conducted, when it was noted that digested sludge dewatered more rapidly than raw sludge. However, design data, are still very empirical, and only recently has an effort been made to develop a rational engineering design approach. An excellent review of past work (1), detailed theoretical analysis, and current understanding of the sludge drying process can be found in various United States Environmental Protection Agency (US EPA) publications (2–4). Sludge dewatering on a drying bed (Figs. 1 and 2) is a multiphase process, which will be amplified in the following sections.

2.1. General Process Description

Sludge drying beds are used to dewater sludge both by draining through the sludge mass and by evaporation from the surface exposed to the air. Usually, collected filtrate is recycled to the headworks in the treatment plant. Drying beds consists of 4–9 in. of sand, which is placed over 8–18 in. of graded gravel or stone. The sand typically has an effective size of 0.3–1.2 mm and a uniformity coefficient of less than 5. Normally, gravel is graded from 1/8 to 1 in. Drying beds have underdrains that are spaced from 8 to 20 ft apart. The underdrainage piping system usually consists of 4-in. pipes, which are made from vitrified clay and laid with open joints at a minimum slope of about 1%.

Sludge is spread on the beds in an 8–12-in. layer. The drying area is partitioned into individual beds, approx 20 ft wide by 20–100 ft long, of a convenient size so that one or two beds will be filled by each sludge withdrawal from the digesters. The interior partitions commonly consist of two or three concerted planks, one on top of the other, to a height of 15–18 in. stretching between slots in precast concrete posts. The outer boundaries for open beds might be of similar construction or earthen embankments. However, concrete foundation walls are required if the beds are to be covered.

Generally, feeder pipes to drying beds are made of cast iron and are designed for a minimum velocity of 2.5 ft/s. The pipes are arranged to drain into the beds and provisions are made to be able to flush the lines and to prevent freezing in cold climates. Distribution boxes are provided to divert sludge flow to the selected beds(s). Splash plates are used at sludge inlets to distribute the sludge over the bed and to prevent erosion of sand.

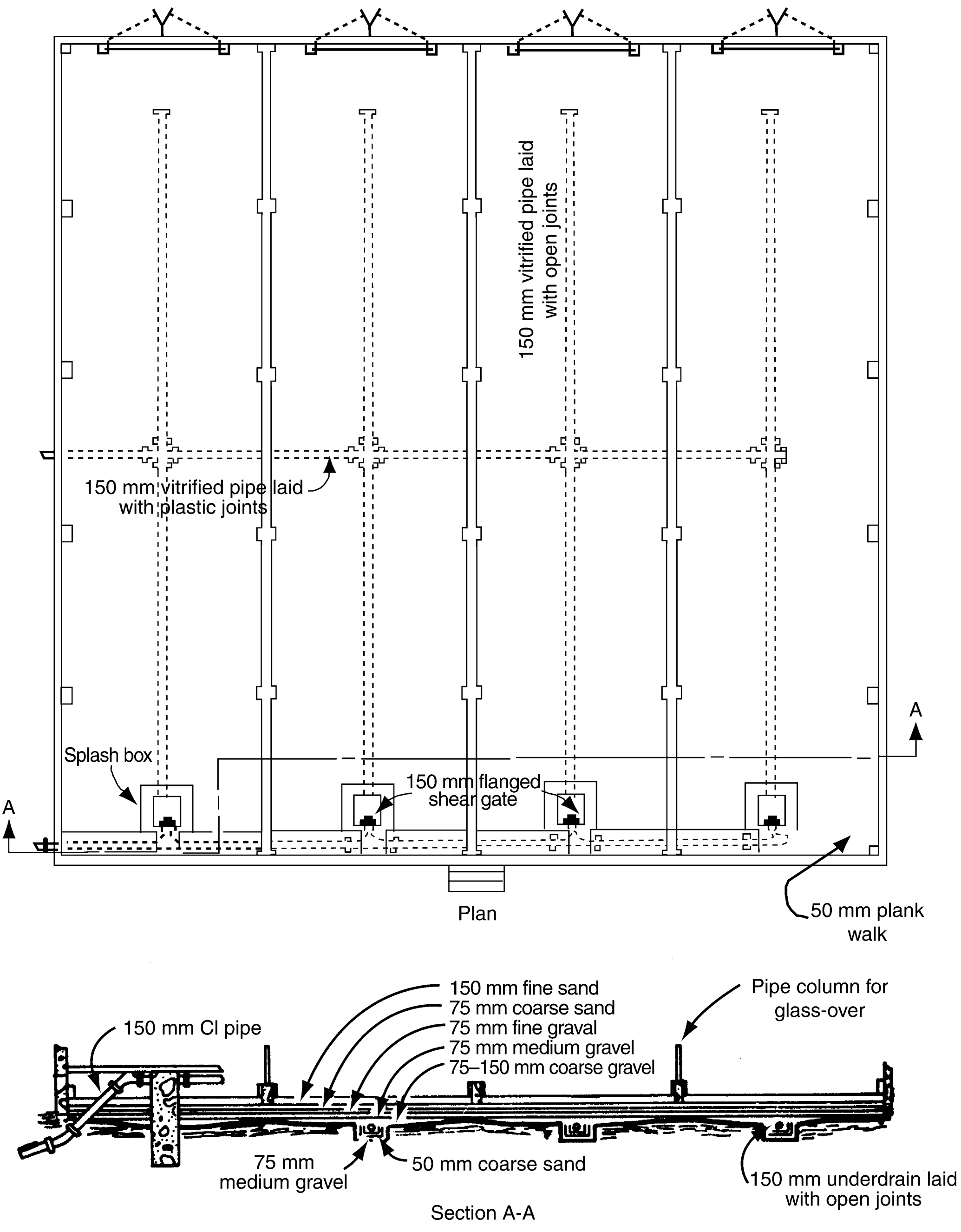


Fig. 2. Sand drying bed construction.

Sludge can be removed from the drying bed after it has drained and dried sufficiently to be separable. Sludge removal is accomplished by manual shoveling into wheelbarrows or trucks or mechanically by scrapper or front-end loader. Provision should be made for driving a truck onto or along the bed to facilitate loading. Mechanical devices can remove sludges of 20–30% solids while cakes of 30–40% concentration generally require manual removal.

Paved drying beds with limited drainage systems permit the use of mechanical equipment for cleaning. Field experience indicates that the use of paved drying beds results in shorter drying times as well as more economical operation when compared with conventional sand drying beds. The reason for this is because of the use of mechanical equipment for cleaning that permits the removal of sludge with higher moisture content than in the case of manual removal. Paved beds have worked successfully with anaerobically digested sludges but are less desirable than sandbeds for aerobically digested activated sludge.

Sand beds can be open or covered depending on climatic conditions. Glass-covered sand beds are protected from rain and cold, improve the appearance of treatment plants, and control odors and insects. To control humidity and optimize water evaporation, the enclosure should be well ventilated. Covered beds require only 2/3–3/4 of the area required for an open bed. However, mechanical sludge removal is more difficult in the former.

2.2. *Operation and Operating Variables*

Dewatering on sand beds proceeds via two different mechanisms; filtration and evaporation. Water drainage is the most important during the first 1–3 d leaving solid concentrations as high as 15–25% (8,9). Further water removal occurs by evaporation. Horizontal shrinking of the sludge and exposure of additional sludge areas facilitate evaporation. It is estimated that 60% of the water is drainable. Up to 85% of the water of secondary sludges can be lost by drainage. In general, the higher the initial water content, the larger the fraction of drainable water (1).

Operational procedures common to all types of drying beds involve:

a. Pump 8–12 in. (20–30 cm) of stabilized liquid sludge onto the drying bed surface.
b. Add chemical conditioners continuously, if conditioners are used, by injection into the sludge as it is pumped onto the bed.
c. When the bed is filled to the desired level, permit the sludge to dry to the desired final solids concentration. This concentration can vary between 18 and 60%, depending on the type of sludge, processing rate needed, degree of dryness required for lifting, and so on.
d. Remove the dewatered sludge either mechanically or manually.
e. Repeat the cycle.

2.3. *Advantages and Disadvantages*

The advantages of using sludge drying beds are as follows:

a. When land is readily available, normally this is the lowest capital cost.
b. Small amount of operator attention and skill is required.
c. Low energy consumption.
d. Less sensitive to sludge variability.
e. Low to no chemical consumption.
f. Higher dry cake solids contents than fully mechanical methods.

On the other hands, the use of sludge drying beds has the following disadvantages:

a. Lack of a rational engineering design approach allowing sound engineering economic analysis.
b. Requires more land than fully mechanical methods.

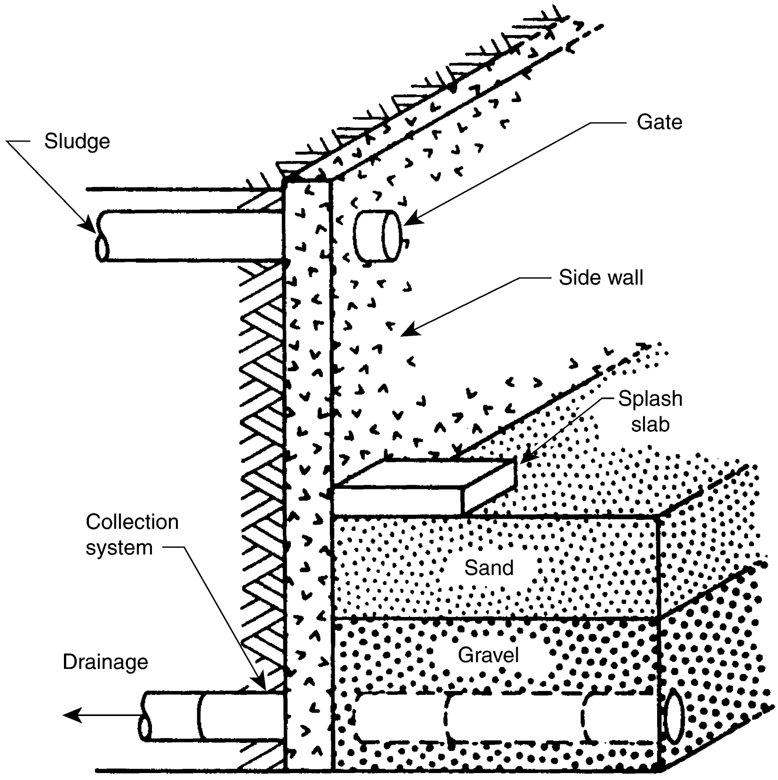


Fig. 3. Typical sand drying bed construction.

c. Requires a stabilized sludge.
d. Must be designed with careful concern for climatic effects.
e. May be more visible to the general public.
f. Removal usually is labor intensive.

3. TYPES OF SLUDGE DRYING BEDS

3.1. Conventional Sand Drying Beds

Sand drying beds are the oldest, the most commonly used type of drying beds. Many design variations are possible including the layout of drainage piping, thickness, and type of gravel and sand layers, and construction materials. Typical sand drying bed construction is shown in Fig. 3. The beds can be built with or without provision for mechanical sludge removal, and with or without a roof.

Current United States practice is to make drying beds rectangular with dimensions of 15–60 ft (4.5–18 m) wide by 50–150 ft (15–47 m) long with vertical side walls. Usually 4–9 in. (10–23 cm) of sand is placed over 8–18 in. (20–46 cm) of graded gravel or stone. Usually, the sand is 0.012–0.05 in. (0.3–1.2 mm) in effective size and has a uniformity coefficient less than 5. Gravel is normally graded from 1/8 to 1 in. (0.3–2.5 cm), in effective size. Normally, underdrain piping has been of vitrified clay, but plastic pipe is also becoming acceptable. The pipes should be no less than 4 in. (100 mm), should be spaced 8–20 ft (2.4–6 m) apart, and have a minimum slope of 1%.

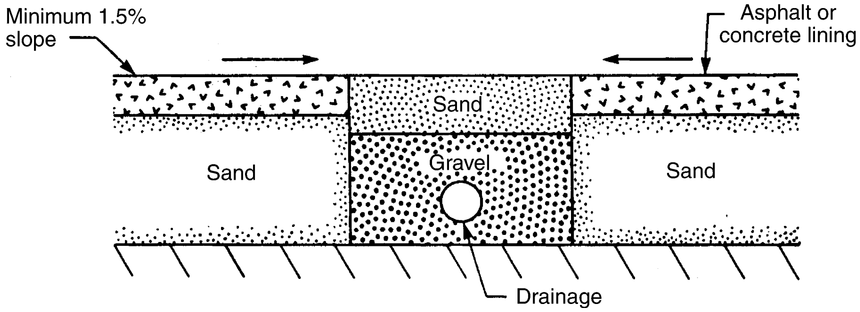


Fig. 4. Typical paved drying bed construction.

3.2. Paved Drying Beds

Since the 1950s, Paved drying beds have had limited use (1). Figure 4 shows typical paved drying bed construction. Normally, the beds are rectangular in shape and are 20–50 ft (5–15 m) wide by 70–150 ft (21–46 m) long with vertical sidewalls. Current practice is to use either concrete or asphalt lining. Normally, the lining rests on an 8–12 in. (20–30 cm) build up sand or gravel base. The lining should have a minimum of 1.5% slope to the drainage area. An unpaved area, 2–3 ft (0.60–1 m) wide is placed along either side or down the middle for drainage. A minimum 4 in. (100 mm) diameter pipe is used to convey drainage water away. Paved drying beds can be built with or without a roof. For a given amount of sludge, paved drying beds requires more area than sand beds. Their main advantages are that front-end loaders can be used for sludge removal and reduced bed maintenance.

3.3. Wedge-Wire Drying Beds

Wedge-wire drying bed systems have been successfully used in England for more than 20 yr to dewater both municipal and industrial wastewater sludge. Used in the United States since the early 1970s, there are presently 18 wedge-wire installations. Ten of these installation are for municipal wastewater sludges. In a wedge-wire drying bed, sludge slurry is introduced onto a horizontal, relatively open-drainage media in a way that yields a clean filtrate and provides a reasonable drainage rate. A typical cross-section of a wedge-wire bed is shown in Fig. 5. The bed consists of a shallow rectangular watertight basin fitted with a false floor of wedgewater panels. These panels have slotted openings of 0.01 in. (0.25 mm). This false floor is made watertight with caulking where the panels abut the walls. An outlet valve to control the rate of drainage is located underneath the false floor.

There are many noticeable advantages of using a wedge-wire drying bed (10):

a. No clogging of the media.
b. Constant and rapid drainage.
c. Higher throughput rate than sand beds.
d. Easy bed maintenance.
e. Difficult-to-dewater sludges, for example, aerobically digested can be dried.
f. Compared with sand beds dewatered sludge is easier to remove.

The procedure used for dewatering sludge begins with the movement of water or plant effluent into the wedge-wire unit until a depth of approx 1 in. (25 mm) over the

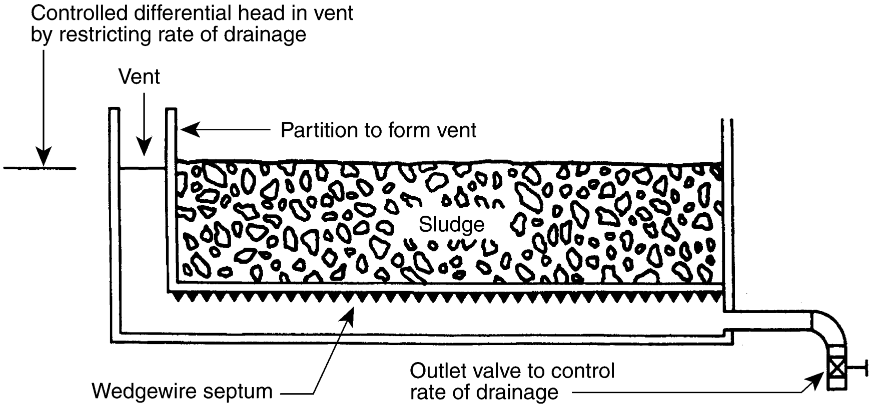


Fig. 5. Cross-section of a wedge-wire drying bed.

wedge-wire septum is attained. This water serves as a cushion that permits the added sludge to float without causing upward or downward pressure across the wedge-wire surface. The water further prevents compression or other disturbance of colloidal particles in the sludge. After the bed is filled with sludge, the initially separate water layer and the drainage water are allowed to percolate away at a controlled rate, through the outlet valve. After the free water has been drained, the sludge further concentrates by drainage and evaporation until there is a requirement for sludge removal.

3.4. Vacuum-Assisted Drying Beds

The only operating vacuum-assisted drying beds at this time are two 20 ft (6m) by 40 ft (12m) units built in 1976 at Sunrise city, Florida. They dewater an aerobically digested sludge having 2% solid concentration, which is wasted from a contact stabilization wastewater treatment plant.

The principal components of the Sunrise facility are as follows:

a. A bottom ground slab consisting of reinforced concrete.
b. A layer of stabilized aggregate several inches thick which provides support for the rigid multi-media filter top. This space is also the vacuum chamber and is connected to a vacuum pump.
c. A rigid multi-media filter top is placed on the aggregate support. Sludge is then applied to the surface of this media.

The operating sequence is as follows:

a. Sludge is introduced onto the filter surface by gravity flow at the rate of 150 gpm (9.4 L/s) and to a depth of 12–18 in. (30–46 cm).
b. Filtrate drains through the multi-media filter and into the space containing the aggregates and then to a sump, from which it is pumped back to the plant by a self-actuated submersible pump.
c. As soon as the entire surface of the multi-media filter is covered with sludge, the vacuum system is started and vacuum is maintained at 1–10 in. Hg (3–34 kN/m^2).

Under favorable weather conditions, this system dewaters the dilute aerobically digested sludge to a 12% solids concentration in 24 h without polymer addition, and to the same level in 8 h if polymer is added. This particular sludge of 12% solids concentration is capable of being lifted from the bed by a fork or mechanical equipment. The sludge will further dewater to about 20% solids concentration in 48 h.

4. PROCESS APPLICATIONS AND LIMITATIONS

There are more than 8000 plants that use open or covered sludge drying beds in the United States. Generally, the sludge drying sand beds are used to dewater sludges in small plants. They require little operator attention or skill. However, air-drying is normally restricted to well-digested or stabilized sludge, because raw sludge is odorous, attracts insects, and does not dry satisfactorily when applied at reasonable depths. Oil and grease clog sandbed pores and thereby seriously retard drainage. The design and use of drying beds are affected by weather conditions, sludge characteristics, land values and proximity of residences. Operation is severely restricted drying periods of prolonged freezing and rain.

5. PROCESS PERFORMANCE, THEORY, AND PRINCIPLES

A cake of 40–45% solids might be achieved in 2–6 wk in good weather and with well-digested secondary, primary, or mixed sludges. With chemical conditioning, dewatering time may be reduced by 50% or more. Solids contents as high 85–90% have been achieved on sand beds, but normally, the times required to achieve such dry sludge cakes are impractical. The mechanisms for water removal impose a number of operating variables that affect the design of drying beds, such as:

a. Sludge condition.
b. Sludge characteristics.
c. Soil permeability.
d. Land availability and cost.

Air-drying of sludge is sensitive to weather conditions. Rain lengthens the drying time, but its effect is less important once shrinkage and cracking has stared (11). However, it reduces the sludge fertilizing value by dissolving valuable nutrients. Air temperature, relative humidity, percentage of sunshine, and wind velocity also affect the rate of water evaporation. In the summer or at high temperature and humidity, the rate of drying is two to three times faster than in the winter or at low temperature (12). It is noteworthy that in many wastewater treatment plants sludge is stored in digesters in the winter and dried only in the summer (11).

Raw sludge does not dry as easily as digested sludge. Its odor and attraction of insects are also serious drawback. Therefore, sand bed drying is usually restricted to well-digested sludges. Entrained gases in digested sludges float solids, leaving a layer of clear water that drains through the sand easily (13). However, Heseltine reported that over digested sludges would also dry slower (14). Primary sludges dry faster than secondary ones. Aged sludges dry slower than new ones. Oil and grease tend to clog the bed pores and retard drying. On the other hand grit facilitates drying (11).

The drying time is affected by the initial concentration of solids in the sludge. Vankleeck reported a doubling of drying time for an increase of solids in sludge from 5 to 8% (15). Laboratory results show that the amount of drainable water decreases approximately linearly with increasing solid concentration, as shown in Fig. 6 (16). Similarly,

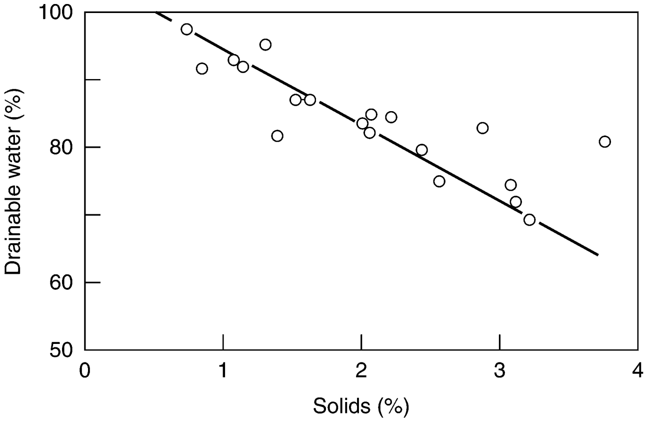


Fig. 6. Effect of solids concentration on drainable water (16).

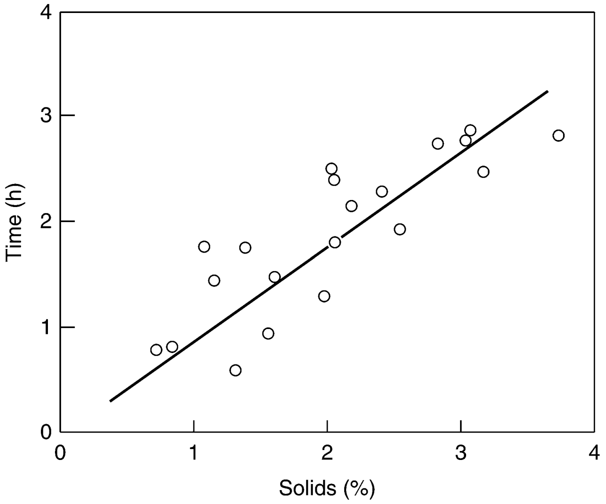


Fig. 7. Dependence of drainage time on solids concentration (16).

the drainage time and the dried cake moisture increase with increasing solid content, as shown in Figs. 7 and 8 (16).

For optimum drying the solids loading of the sand bed should not exceed 15 lb of dry solid/ft^2 for uncovered beds and 25 lb/ft^2 for covered beds (17). The rate of drying depends on the depth of the sludge charge over the sand bed (18). Drainage time and cake moisture increase with increasing depth of the charge (12,16), as shown in Figs. 9 and 10.

The drying rate can be increased by sludge conditioning with organic or inorganic coagulants-flocculents. Chemical conditioners permit higher sludge porosity, decrease solids compression, and result in reduced sand bed maintenance (11). Inorganic chemicals such as alum, ferric chloride, sulfuric acid, anthracite, and activated carbon have been used (19). Several organic polyelectrolytes have been tested (19,20) sometimes

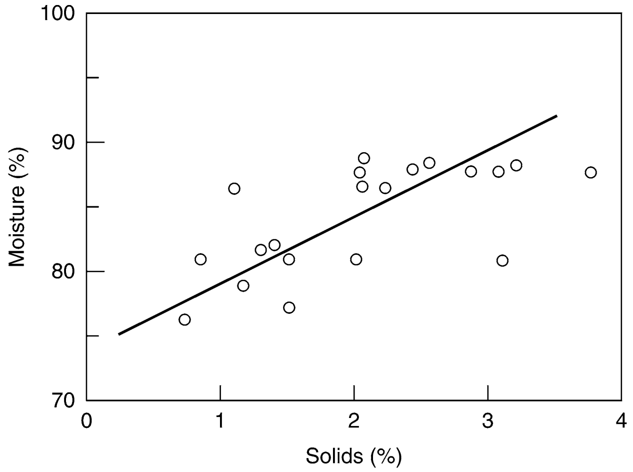


Fig. 8. Cake moisture dependence on solids concentration (16).

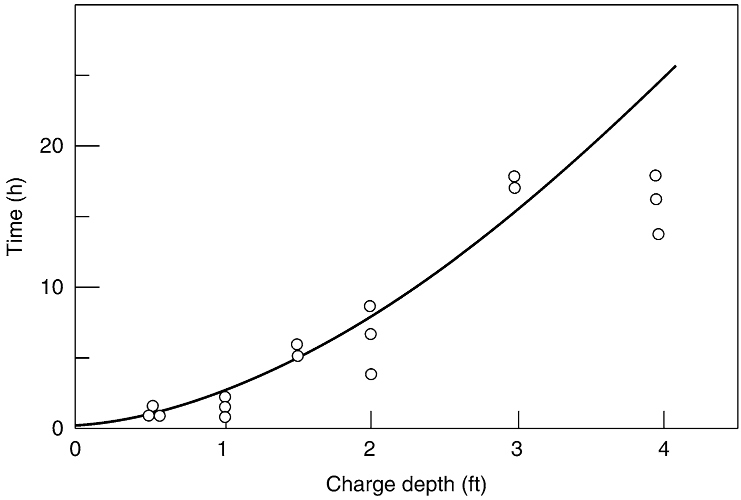


Fig. 9. Drainage time dependence on initial depth of sludge change (16).

with conflicting results. Sperry found alum the most effective and economic conditioning additive (21). Alum, aluminum chlorohydrate, and ferrous sulfate are in common use in Britain (22). Recommended alum dosage is 1 lb/100 lb of sludge (15,23). For thin sludges (<3% solids), one-half to one-third of this dosage suffices (21). Ferric chloride is the most commonly used inorganic coagulant in the United States. A dosage of 90 lb/t (dry) results in a liftable dried sludge after 10–20 h (24).

The amount of inorganic ionic coagulants required for effective sludge drying depends on the sludge pH. Because metal ion conditioning additives lower the pH, highly buffered sludges require large amount of coagulants before any noticeable pH change occurs. Weber (19) discusses in detail the reactions, the effect, and the mode of operation of ionic coagulants.

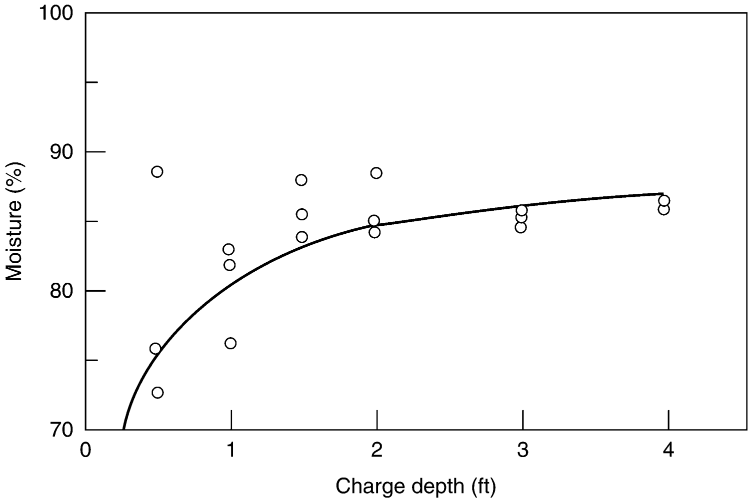


Fig. 10. Effect of initial change depth on cake moisture content (20).

In contrast to the high dosages of inorganic additives, organic polyelectrolytes are effective at dosages less than 1% of the solids weight. However, they are appreciably more expensive than the inorganic salts. The efficiency of organic polymers in promoting drying decreases in the order (25): Raw primary sludge > digested primary sludge > digested activated sludge. Inorganic sludges are less susceptible to conditioning with organic polyelectrolytes. Some improvement with certain polymers was reported (26). Neubauer observed no effect on inorganic sludge drying (10). Organic flocculants do not only increase the rate of drying by as much as 30 times, but they also permit heavy bed loading up to 3 ft (27). Figures 11 and 12 show the effect of polymer dosage and bed depth on the solids concentration with time.

Most of the discussion above referred to conventional beds, built directly on soil. Attempts to use mechanical cleaning of sand beds or to increase drying rates, led to alternative bed designs such as paved beds, wedge-wire beds, and heated beds. Evaporation becomes important with some of these bed designs. Asphalt and concrete paved beds proved that bed performance was not impaired by the pavement (28–32). Shorter drying times were reported, because mechanical lifting permitted removal of sludges with higher water content compared with manual cleaning (29).

Wedge-wire beds have operated successfully in England (11,33,34) and in the United States (22). A diagram of wedge-wire bed is shown in Fig. 5. Initially, support water is introduced to prevent blinding of the filtering medium. After the sludge is applied to the bed, this free water is drained and the sludge is allowed to dry by drainage and evaporation. The following difficult sludges have been dewatered successfully over wedge-wire beds: tannery sludges, slag fines, bacterial slimes, vegetable waste, and hydroxide sludges (33–35). Wedge-wire beds appear to be resistant to clogging, less susceptible to adverse weather, easy to clean and maintain. Bed capacity is increased because of shorter drying and cleaning times. Table 1 illustrates the performance of wedge-wire beds as a function of sludge type, solids concentration, and dewatering time.

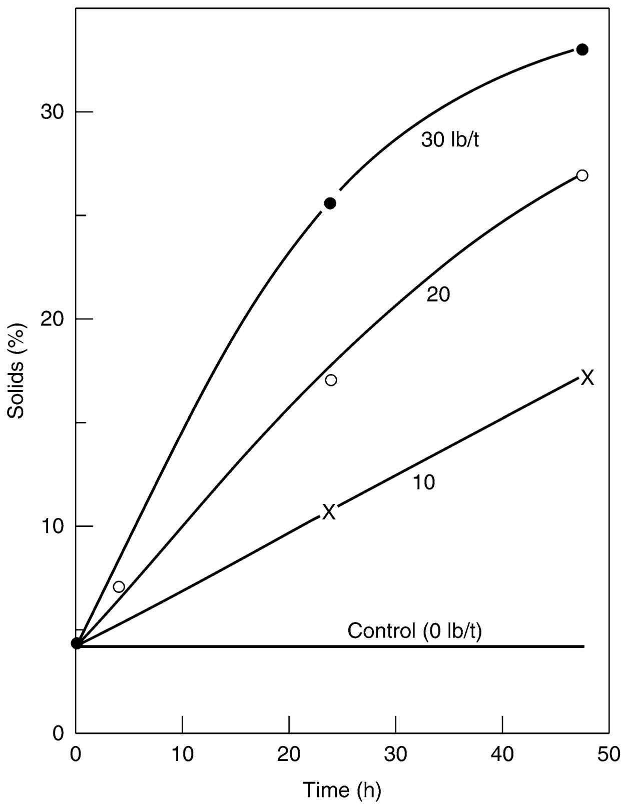


Fig. 11. Dewatering of digested sludge treated with varying amounts of polymeric flocculant (27). Reprinted with permission from the Dow Chemical Company.

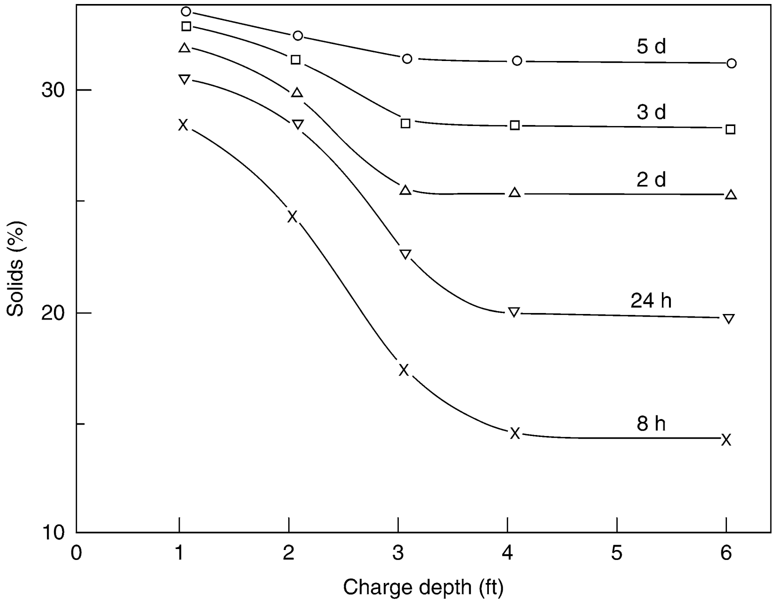


Fig. 12. Effect of initial depth on solids concentration in the cake for an 8% sludge treated with a polymeric flocculant dose of 60 lb/t (2). Reprint with permission from the Dow Chemical Company.

Table 1
Wedge-Wire System Performance Data

Sludge type[a]	Feed solid (%)	Sludge solid concentration (%)	Dewatering time	Solid capture (%)
Primary	8.5	25	14 d	99
Trickling filter humus	2.9	8.8	20 h	85
Digested primary + WAS	3	10	12 d	86
Fresh WAS	0.7	6.2	12 h	94
Fresh WAS	1.1	9.9	8 d	87
Thickened WAS	2.5	8.1	41 h	100

Source: From ref. 34.
[a] All sludges were chemically conditioned.
WAS, waste-activated sludge.

Table 2
Recommended "Ten States Standards" Sand Bed Areas for 40 and 45°N Latitude

	Bed area (ft^2/capita)	
Type of sludge	Open bed	Covered bed
Primary	1	0.75
High rate trickling filter	1.50	1.25
Activated sludge	1.75	1.35

Source: From ref. 12.

Heating sludge with hot water, circulated into covered beds through heating coils, improved the bed drying capacity appreciably (11,36). Heating probably accelerates biological decomposition to form gases that float the solid particles and allow water drainage. It also breaks down the colloidal structure of the sludge and promotes particle coalescence (37). Finally, heating decreases the sludge viscosity, thereby improving drainage rates.

6. DESIGN CRITERIA, REQUIREMENTS, AND OTHER CONSIDERATIONS

6.1. Land Requirements

The rate of sludge drying, depending on the previously discussed parameters determines the minimal land areas required for beds installation. Recommended sizes for drying sand beds are given in Table 2. Open beds require larger areas than covered beds. In warmer climates, area requirements may be lower than those shown in Table 2. Furman recommends an area of only 0.3 ft^2/capita for drying of digested primary sludge and of trickling filter humus in Florida (38). This is less than 1/3 the required size in the northern states.

Eckenfelder and O'Connor recommend annual solid loadings per unit bed area for various types of wastewater sludges in the range of 15–27 lb/ft^2 as given in Table 3 (8). The effect of sludge type on recommended bed area requirements in England is shown in Table 4.

The area requirements decrease for chemically conditioned sledges. This is because of the higher solids loadings permitted for conditioned sludges, the application of deeper

Table 3
Solids Loading Per Unit Bed Area

Type of sludge	Sludge loading rate (dry solids lb/ft^{2}yr)
Primary	27.5
Primary trickling filter	22
Primary activated	15
Chemically precipitated	22

Source: From ref. 8.

Table 4
Recommended Sand Areas for England

Type of sludge	Open bed area (ft^2/capita)
Primary	1.3
Trickling filter	1.5
Digested, mixed	1
Undigested, mixed	2.25
Greasy sludge	3

Source: From ref. 39.

sludge loads, and the faster drying rates of such sludges. Land unavailability limits the application of sludge drying sand beds in the vicinity of large cities. However, sand beds will continue to be important sludge drying facilities in medium and small cities and towns. Pipeline transportation of fluid sludges to rural areas might prove economical and advantages for sludge drying and land application.

6.2. Covered Beds

Whenever there is the possibility of long period of rain, snow or cold weather, potential odor or insect problems, or a problem with esthetics, consideration should be given to employing covers for the drying beds. When properly ventilated, so that air can flow over the surface of the beds, covered sand beds can be used and require 25–33% less area than open sand beds (*see* Table 2). Although covers can be provided for paved, wedge-wire, and vacuums beds, no documentation is available on how covers would affect or improve loading rates for such beds.

6.3. Sludge Conditioning

Sludge conditioning can dramatically improve drying bed throughput and should be taken into consideration as part of the design. This is an especially important consideration in the case of difficult to dewater sludges.

6.4. Sludge Removal

The majority of United States facilities use manual labor to remove dried sludge from drying beds. With this type of removal, a 30–40% solid concentration is required. With mechanical sludge removal systems, solids concentration between 20 and 30% can be handled. Depending on the bed size, a tillable unit similar to the lift and dump mechanism of a dump truck is available for the wedge-wire drying bed.

6.5. Sidestreams

The only Sidestream for a drying bed operation is the under drainage liquor. Normally, this flow is not treated separately, but is typically recycled to the plant headworks. The effect of the high nitrogenous content of this recycled stream should be taken into consideration in the design of the upstream units in the treatment plant.

The following are the characterization results from a sand bed drainage study in the United States:

a. Sludge type = anaerobically digested mixture of primary and trickling filter sludge.
b. Bed media = 6 in. of sand.
c. Color = clear, dark amber.
d. COD = 300–400 mg/L.
e. BOD_5 = 6–66 mg/L.
f. BOD_{20} = 1900–2360 mg/L (>90% nitrogenous).

6.6. Bed Sizing Criteria

The majority of published and professionally utilized design data are based on operations during the 1940s and 1950s. Tables 5 and 6 summarize the available data for the sizing of sand drying beds. At that time, sludges applied to sand beds were predominantly anaerobically digested. They originated from plants having primary, primary plus trickling filters, or primary plus conventional activated sludge treatment processes. Many of the sludges presently generated do not readily fall within these categories.

It should also be noticed that most data are reported in terms of square feet of bed surface area required for dewatering on a per capita basis (Tables 5 and 6). This criterion is only valid for the characteristics of a particular wastewater and has no rational design basis. The better criterion for sizing sand drying beds is in terms of solids loading, pounds of solids per square foot of bed surface area per year (Table 5). The best criteria would also take into consideration the climatic conditions (such as temperature, wind velocity, and precipitation), sludge characteristics (grit, grease, fiber, and biological content), and solids concentration.

No generalized bed sizing criteria could be found for paved beds. Table 1, which was referred to earlier, contains the performance data for wedge-wire systems. In addition to the solids concentration, in feed and dried sludges, and the dewatering time for various types of sludges, the table also reports that the solids capture ranges between 85 and 100%.

7. ENVIRONMENTAL IMPACT AND ENERGY CONSUMPTION

Land requirements are large. Odor can be a problem with poorly digested sludges and inadequate buffer zone areas. Total annual energy consumption is the summation of mechanical scraping energy needs, sand replacement energy consumption and pumping energy requirement:

$$E_t = E_p + E_{ms} + E_{st} \tag{1}$$

Table 5
Summary of Recognized Published Sand Bed Sizing Criteria for Anaerobically Digested, Nonconditioned Sludge

	Uncovered bed		Covered bed
Initial sludge source	Area (ft^2/capita)	Loading (lb solids/ft^2/yr)	Area (ft^2/capita)[a]
Primary			
Ref. 51	1	27.5	
Ref. 52	1–1.5		0.75–1
Ref. 53			
N45°N latitude	1.25		0.93
Between 40–45°N	1		0.75
S40°N latitude	0.75		0.56
Primary + chemicals			
Ref. 51	2	22	
Ref. 52	2–2.25		1–1.25
Ref. 53			
N45°N latitude	2.5		1.87
Between 40–45°N	2		1.5
S40°N latitude	1.5		1.12
Primary + low rate tricking filter			
Ref. 51	1.6	22	
Ref. 52	1.25–1.75		1–1.25
Ref. 53			
N45°N latitude	1.87		1.56
Between 40–45°N	1.5		1.25
S40°N latitude	1.12		0.93
Primary + WAS			
Ref. 51	3	15	
Ref. 51	1.75–2.5		1.25–1.5
Ref. 53			
N45°N latitude	2.18		1.68
Between 40–45°N	1.75		1.35
S40°N latitude	1.31		1.01

[a] Only area loading rates available for covered beds.
1 lb/ft^2/yr = 4.9 kg/m^2/yr.
1 ft^2 = 0.093 m^2.

$$E_p = 3.89 \times 10^6\, Q\, (\text{TDH}) \times 100/\varepsilon \tag{2}$$

$$E_{ms} = \text{estimated to be } 3.2 \times 10^6 \text{ Btu/yr/MGD at 900 lb dry solids/MG} \tag{3}$$

$$E_{sr} = \text{estimated to be 10\% of } E_{ms} = 0.32 \times 10^6 \text{ Btu/MGD} \tag{4}$$

where E_t is the total annual energy consumption (Btu/yr); E_{ms} is the mechanical scraping energy consumption (Btu/yr); E_{sr} is the sand replacement energy consumption (Btu/yr); E_p is the pumping energy consumption (Btu/yr); ε is the wire to water efficiency (%); Q is the sludge flow (MGD); TDH is the total dynamic head (ft).

Table 6
Summary of Recognized Published State Bed Sizing Criteria for Sand Beds by US EPA Regions (ft^2/capita)[a]

	EPA region												
	I		II		IV[b]	VI		VII[b]	VIII		IX[b]	X[c]	
	U[d]	C[d]	U	C	U	U	C	U	U	C	U	U	C
Anaerobically digested primary only	1.5	1	1.5	0.75	0.5–1	1		1				1.5	1
Primary + low rate trickling filter	1.75	1.25	1.5	0.75	0.75–1.2	0.5–1	0.25	1.5	1	1	1	1.5–2	1–1.25
Primary + sand filter					1	1					0.5		
Primary + high rate trickling filter					1	1			1.25	1.25	1	2	1.25
Primary + WAS	2.5	1.5	2	1	1.5–2.5	1–1.5	1		1.35	1.35	1	1.5–2.5	1–1.5
Primary + chemical			2	1	1–1.33	1			1.5	1.3		3	2
Imhoff			1.5	0.75	0.66–1	1							
Imhoff + low rate trickling filter					1–1.2	1							

[a] Taken from individual state design criteria that do not use 10 States standards.

[b] The state encompassed in US EPA regions IV,VII, and IX do not have published requirements for covered sand beds.

The state encompassed in US EPA regions III and V do not have published requirements at this time.

[c] State of Idaho: values shown are for rainfall of 30–45 in. (76–114 cm): for rainfall between (10 and 30 in. [25–76 cm]), reduce those values by 50%.

[d] U, uncovered sand beds; C, covered sand beds.

Table 7
Sludge Drying Beds, Labor Requirements[a]

Total bed area[b] (ft^2)	Labor (h/yr)		
	Operation	Maintenance	Total
1000	300	100	400
5000	400	180	580
10,000	500	220	720
50,000	1500	710	2210
100,000	2900	1500	4400

[a] From ref. 7.

[b] Assumes dry solid loading rate of 20 lb/ft^2/yr of bed area.

1 ft^2 = 0.093 m^2.

1 lb/ft^2/yr = 4.9 kg/m^2/yr.

8. COST

8.1. Capital Costs

US EPA and other researchers have developed capital cost curves for open sand beds (40–42). The authors have updated the literature data to January 2002 (ENR CC Index = 6390.21). Although the data were scattered, a regression analysis indicated that, on the basis of a US EPA Municipal Wastewater Treatment Plant Construction Cost Index for the 2nd quarter of 1977, the capital construction cost could be approximated by Eq. (5):

$$C_c = 25.27 \times 10^4 \, Q^{1.35} \tag{5}$$

where C_c is the capital construction cost of sand beds (USD); Q is the plant design wastewater flow (MGD).

The associated costs include excavation, process piping, equipment, concrete, and steel. In addition, such costs, as those for administration and engineering could be calculated from Eq. (6)

$$C_{ae} = 0.2264 \times C_c \tag{6}$$

Hence, the total capital cost is:

$$C_t = C_c + C_{ae} \tag{7}$$

where C_{ae} is the cost for administration and engineering; C_t is the total capital cost.

8.2. Operating and Maintenance Cost

The labor requirements as a function of open sand bed surface area for both operation and maintenance are given in Table 7. The labor indicated includes: removal of dried sludge from the beds, sand maintenance, and weeding as necessary.

The maintenance material costs as a function of sand bed surface area are shown in Table 8. As an example, for a sand bed having a surface area of 10,000 ft^2 (930 m^2), a designer would estimates a yearly materials cost of 1022 USD. The authors have adjusted the cost data to the design period of January 2002.

Table 8
Estimated Maintenance Material Cost for Open Sand Drying Beds[a]

Drying bed area (ft^2)	Annual cost (USD)
5000	562
10,000	1022
50,000	5368
100,000	9713
500,000	48,566

[a] January 2002 cost data (ENR CC Index = 6390).
1 ft^2 = 0.093 m^2.

Table 9
Process Variables, Measurements, and Instruments of Sludge Drying Beds

Process variables	Measurements	Instruments
Feed sludge	Flow	Venturi with diaphragm sensors Magnetic Doppler Pump displacement
	Pressure	Bourdon with cylindrical seal
	Density	Nuclear Optical Ultrasonic
Bed contents	Moisture contents	Portable ohmmeter Lab test
Dewatered sludge	Flow	Transport displacement
	Weight	Static
	Moisture content	Portable ohmmeter Lab test
Weather	Wind speed (15 ft [4.6 m] above ground)	Anamometer
	Wind direction (15 ft [4.6 m] above ground)	Wind vane
	Temperature (dry bulb 5 and 25 ft [1.5 and 7.6 m] above ground)	RTD with solar shield Thermistor with solar shield
	Relative humidity	RTD with lithium chloride cloth (wet bulb temperature)
	Rainfall	Tipping bucket
	Solar radiation	Thermopile
Atmospheric monitoring	Odors	Portable olefactometer

Source: US EPA.

9. PROCESS MONITORING

The salient process monitoring variables, required measurements and recommended monitoring instruments have been established by the US EPA. Table 9 summarizes what measurements and instruments are needed for feed sludge, bed contents, dewatered sludge, drainage and surface runoff, weather, and atmosphere.

Table 10
Summary of Septage Dewatering Studies Using Sludge Drying Beds

Chemical addition	Remarks	References
Alum	Alum-enhanced dewatering	45
Alum, ferric chloride, lime and acid	Screened raw septage dewatered to 6%; $FeCl_3$/lime dewatered to 11% cake solids; alum-treated dewatered to 15%; And acid/lime conditioned septage dewatered to 24% cake solids in 2 d	42
Alum and aluminum potassium sulfate	Conditioning septage to a CST of <50 s enabled dewatering in 48 h with cake solids of 20%	43
Lime	Lime added to pH>10.0, septage then dewatered to 25% cake solids in 6 d	40

10. DESIGN AND APPLICATION EXAMPLES

10.1. Example 1

One of the important applications of sludge drying beds is dewatering of septage. Please introduce this application.

Solution

Dewatering of septage using sand drying beds is a convenient method for small communities in areas where land availability is not a major constraint. Septage is placed on drying beds of sand and gravel and allowed to dry. Dewatering occurs by drainage through the sand bed and also by evaporation. As the septage dries, cracks develop on the surface allowing further evaporation and drying in the lower layers. The filtrate draining through the sand is collected in a perforated, open-jointed piping system below the sand beds. It can either be returned to the headworks of the treatment plant or treated separately as an effluent before ultimate disposal. The sludge cake is removed from the sand bed either by front loaders or by hand shoveling, and then it is trucked for land application or to an ultimate disposal site. A typical sand bed drying system is shown in Figs. 1 and 2.

Table 10 shows various applications of sludge drying beds for dewatering septage. The given examples illustrate the various combinations of drying times and conditioning chemicals that can be used to produce a variety of cake solids concentrations. Feige (43) noted that the addition of approx 90 kg lime/T (180 lb lime/t) of dry septage solids resulted in 25% cake solids in 6 d and 38% in 19 d. Condren's studies (44,45) showed that alum-conditioned septage dewatered to 15% cake solids after 1 d, whereas ferric chloride/lime-conditioned septage produced 10–11% cake solids after 2 d. In comparison, acid lime conditioning of septage resulted in a cake of 24% solids after 2 d. Perrin (46) evaluated dewatering characteristics of septage in laboratory-scale studies using capillary suction time (CST) as the parameter for comparison. Perrin found that septage with a CST of 50 s would cease free drainage on a sand drying bed within 48 h or less, resulting in about 20% cake solids content. Studies by

Shaboo and Crowe (47,48) indicated that a CST of 50 s can be achieved by conditioning septage with 0.1–0.2 kg lime/kg dry solids. However, there is no basis to compare these two CST values as a result of differences in total solids and test methods.

Because evaporation is a contributing factor to the performance of sand bed dewatering, depth of application of septage is an important design/operation consideration. One study indicated that chemically-conditioned septage dewatered more readily at 15 cm (6 in.) depth of application than at a 30 cm (12 in.) depth (48). Based on this and other pilot-scale and full-scale dewatering plants, a septage application depth of 20 cm (8 in.) is recommended.

Drying time is a function of drainage and evaporation. The predominant function of sand bed dewatering is drainage, most of which usually occurs within about 7–10 d. However depending on weather conditions, evaporation also contributes significantly to dewatering, particularly in the latter part of the drying period. The average drying time is about 2–4 wk. Because conditioning significantly improves dewatering characteristics of septage, it is possible that the average drying time for conditioned septage might be reduced to approx 10–15 d. Table 11 provides a summary of findings of some studies on sand bed dewatering of septage. Although bench scale and pilot plant studies indicate drying time for septage between 2 and 6 d, full-scale operations are estimated to require longer drying time.

Sand bed drying is one of the simplest systems that can be used for dewatering of conditioned septage. The advantages of this system are:

a. Simple construction.
b. Minimal operator training and attention required.
c. Low capital and operation costs.

The disadvantages are as follows:

a. Large sand area required.
b. Potential problems with operation during cold and wet weather seasons unless the beds are covered.

One of the variations in sand drying bed construction relates to the choice between asphalt and concrete paved drying beds. Use of mechanical equipment for cleaning unpaved sand beds has resulted in damage to underdrain pipes. Paved drying beds permit the use of mechanical equipment without damaging underdrains and thereby reduce the cost of labor and sand replacement. Usually Paved drying beds are constructed with a 1.5–2% slope toward the center. A perforated drainage pipe is located in the center beneath a sand drainage strip at a level below the paved bed. Operation of paved drying beds is economical because the use of mechanical equipment allows removal of sludge with higher moisture content in shorter drying time intervals than in the case of manual cleaning. The main disadvantage of paved beds is higher capital cost. The feasibility of using paved drying beds for dewatering digested sludge has been demonstrated elsewhere (49–51,56).

10.2. Example 2

Introduce the newly developed vacuum-assisted drying bed and its application for septage disposal.

Table 11
Summary of Studies on Sand Bed Dewatering of Septage

Type of study (Reference)	Feed septage (mg/L)	Conditioning		Drying time (d)	Cake solids (mg/L)	Filtrate (mg/L)	Comments
		Chemical	Dosage (mg/L)				
Pilot study (42)	TSS-7700	None	–	2	TSS-59,500	TSS-319	0.2 m^3 batches were placed on 1 m^2 sand beds
	TSS-21,000	$FeCl_3$ $Ca(OH)_2$	400 4000	2	TSS-105,000	TSS-46	Conditioned septage was settled for 22 h
	TSS-30,600	Alum	355–955	1	TSS–153,000	TSS-79	0.2 m^3 screened septage was placed in 20 cm layers on a 30.5 cm deep bed with an area 0.93 m^2
	TSS-21,000	(H_2SO_4) $Ca(OH)_2$	3000–4000	2	TSS-241,000	TSS-53	$Ca(OH)_2$ dosage to raise pH to 11.5, an
Pilot study (40)	TS-37,000 COD-58,000	$Ca(OH)_2$	83,000	6 19	TS-247,500 TS-380,000	– COD-186-1660	average of four tests sand beds of 6 m^2 area covered with septage to a depth of 20 cm
Bench scale (43)	TSS-32,000	Aerated	–	3–4	TS-200,000	–	Added 0.5 L of septage to sand drying column
		$FeCl_3$	640–1280	3–4	TS-200,000		
		Alum	70–135	3–4	TS-200,000		
		Polymer	640–1280	3–4	TS-200,000		
Bench scale (45)	TS-34,500 VS-14,300 COD-17,000	Alum	100	1.5	TS-150,000	– – –	

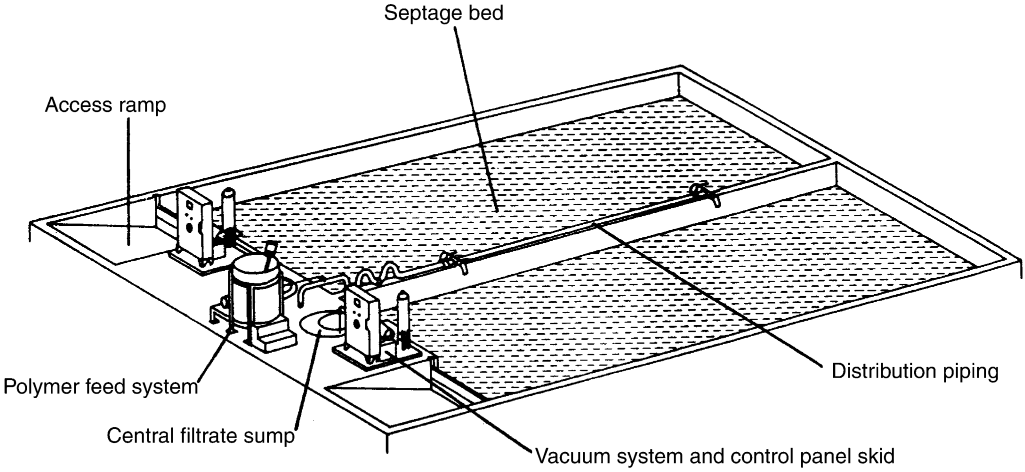


Fig. 13. Vacuum assisted drying bed system (5,52).

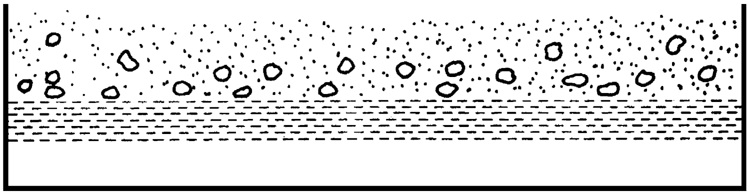
Solution

This is a relatively new system for dewatering chemical sludges. It consists of a drying bed of permeable media to which polymer-treated sludge is applied to depths of about 30 cm (1 ft), and allowed to drain by gravity. Vacuum is then applied and held until the sludge surface cracks. The sludge cake is then ready for removal by front-end loaders specially equipped with rubber-bottomed buckets. After a washing step, the process can again be initialed. The process is schematically illustrated in Fig. 13. Proprietary systems of three different manufacturers are available. These systems are quite similar and are designed on the concept that a vacuum applied to a permeable mat loaded with sludge significantly improves the dewatering efficiency. Figure 14 outlines the steps involved in the operation of a typical vacuum-aided drying bed.

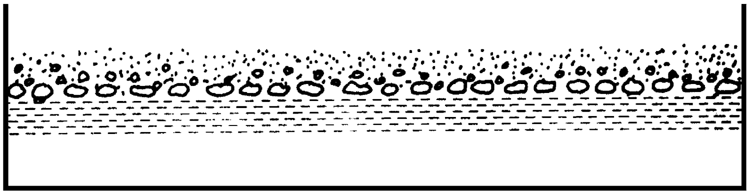
This system of dewatering septage may have several advantages:

a. Simplicity in construction.
b. Minimal operator training and attention.
c. Able to produce truckable sludge cake in 24 h.
d. Generally porous blocks resist clogging and blinding and require only hosing down for cleaning.
e. High loading rate capability.

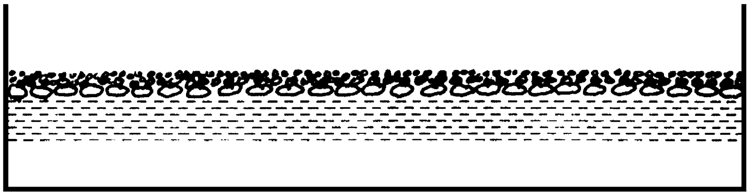
Although open-air drying beds may require large areas, the provision of vacuum significantly reduces the area requirement over that for conventional sand drying beds. Moreover, the ability to remove sludge cake in 24-h or less cycles reduces the total bed area requirements. As with other dewatering systems for septage, chemical conditioning is required before application on the bed. The level of septage conditioning required is not yet tested, but experience with sewage treatment plant (STP) sludge would indicate that lime stabilization or aerobic or anaerobic digestion would suffice before polymer treatment. Cake sludge solids ranging between 15 and 25% are claimed to be achieved in 24 h or less, which is a very high yield compared with gravity sand drying beds (5,52,56).



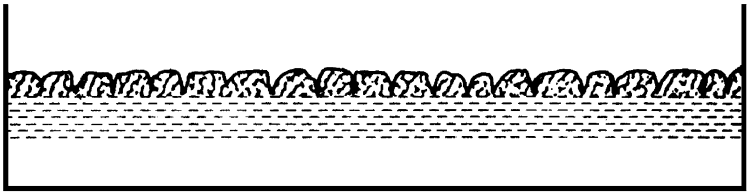
① Conditioned sludge is distributed on the permeable mat. Immediately large particles begin to settle onto the mat as free water drains.



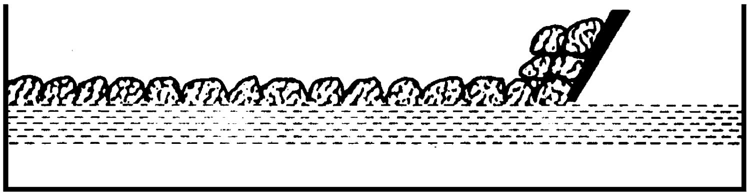
② A cohesive layer of large particles formed on the mat traps finer sludge particles as free water continues to drain.



③ A vacuum is applied to the bed drawing additional free water trapped between the sludge particles.



④ The vacuum continues dewatering the sludge until it cracks, allowing passage of air through the bed, continuing the dewatering process.



⑤ Once the sludge has been dewatered to a truckable condition it is removed from the mat.

Fig. 14. Operational sequence in vacuum assisted sludge drying bed.

This system of dewatering appears to be well-suited for independent septage treatment as it combines the simplicity of a gravity dewatering system with the rapid dewatering rates of a mechanical system. As in many cases, independent septage treatment systems might be small in size and might be located in relatively remote areas, a simple, efficient system with low maintenance requirements is highly desirable. Depending on the pattern of septage generation and climatic conditions, generally these beds will require heated enclosures in northern regions. No performance data on application of this system to septage treatment is currently available. Conventional sand drying beds may be loaded from 10 to 270 kg dry solids/m^2/yr (2–55 lb/ft^2/yr) depending on type of sludge, weather, dryness required, and whether the bed is covered or uncovered. Loading for operating vacuum assisted sludge drying beds has ranged up to 950 kg dry solids/m^2/yr (195 lb/ft^2/yr), with typical polymer additions between 2 and 6 kg/T of dry solids (4 and 12 lb/t). Manufacturers claim that significantly higher loading is possible (53). Here 1T = 1000 kg and 1t = 2000 lb.

10.3. Example 3

There are no prefabricated sludge drying beds available commercially. Introduce the engineering consultants who are qualified for sludge drying beds design, and might be retained for their services (54–60).

Solution

In addition to the authors of this chapter, other qualified design engineers might be found from literature.

10.4. Example 4

A wastewater treatment plant has a drying bed 200 ft long and 20 ft wide. The sludge generated at the plant has a solids concentration of 5%. Sludge is applied to the bed up to a maximum depth of 9 in. Calculate the number of pounds of solids that can be dried each year for every square foot of drying bed area. Assume that it takes 3 wk for sludge drying and 2 d for solids removal (60).

Solution

Volume of applied sludge = 200 × 20 × (9/12) ft^3 = 3000 ft^3
= 3000 ft^3 × (7.48 gal/ft^3) = 22,440 gal.
Weight of dry solids (lb) = 22,440 × (8.34 lb/gal) × (5/100) = 9360 lb.
Number of days/cycle (drying + removal) = 3 wk × (7 d/wk) + 2 d = 23 d
Daily weight of dry solids production = 9360/23 = 407 lb/d.
Yearly weight of dry solids produced = 407 lb/d × 365 d/yr. = 148,600 lb/yr
Therefore, the weight of solids that can be dried each year for every square foot of drying bed area is, (148,600 lb/yr)/(200 ft × 20 ft) = 37 lb/yr/ft^2.

10.5. Example 5

A wastewater treatment plant has a total sludge production of 0.5 MGD, which is pumped to the drying beds at a total dynamic head of 40 ft with wire to water efficiency of 60%. Determine the total annual energy consumption.

Solution

From Eq. (2), the pumping energy requirement is:

$$E_p = 3.89 \times 10^6\, Q\, (\mathrm{TDH}) \times 100/\varepsilon \qquad (2)$$

$$= 3.89 \times 10^6 \times 0.50 \times 40 \times 100/60$$

$$= 129.7 \times 10^6 \text{ Btu/yr.}$$

From Eq. (3), the mechanical scraping energy needs are:

$$E_{ms} = (3.2 \times 10^6 \text{ Btu/yr/MGD}) \times (\text{Q MGD}) \qquad (3)$$

$$= 3.2 \times 10^6 \times 0.5 = 1.6 \times 10^6 \text{ Btu/yr.}$$

From Eq. (4), the sand replacement energy consumption is:

$$E_{sr} = 10\% \text{ of } E_{ms} \qquad (4)$$

$$E_{sr} = 0.10 \times 1.6 \times 10^6 = 0.16 \times 10^6 \text{ Btu/yr.}$$

From Eq. (1), the total annual energy consumption in Btu/yr is:

$$E_t = E_p + E_{ms} + E_{sr} \qquad (1)$$

$$E_t = 129.7 \times 10^6 + 1.6 \times 10^6 + 0.16 \times 10^6 = 131.46 \times 10^6 \text{ Btu/yr.}$$

Hence, the total annual energy consumption E_T in kWh/yr is:

$$E_T = 131.46 \times 10^6 \text{ Btu/yr} \times 2.928 \times 10^{-4} \text{ kWh/Btu}$$

$$= 38{,}500 \text{ kWh/yr.}$$

NOMENCLATURE

C_{ae} = cost for administration and engineering, USD
C_c = capital construction cost of sand beds, USD
C_t = total capital cost, USD
E_{ms} = mechanical scraping energy consumption, Btu/yr
E_p = pumping energy consumption, Btu/yr
E_n = sand replacement energy consumption, Btu/yr
E_t = total annual energy consumption, Btu/yr
E_T = total annual energy consumption, kWh/yr
Q = flow, MGD
t = ton (English ton)
T = Tonne (Metric ton)
TDH = total dynamic head, ft
ε = wire to water efficiency, %

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14

Animal Wastes Treatment Using Anaerobic Lagoons

Lawrence K. Wang, Yung-Tse Hung, and J. Paul Chen

CONTENTS

1. INTRODUCTION

In many situations, it is necessary to pretreat agricultural waste before final treatment. The purpose of pretreatment is to reduce pollution potential of the waste through biological, physical, and chemical processes. These types of components reduce nutrients, destroy pathogens, and reduce total solids. Pretreatment also includes solids separation, drying, and dilution that prepare the waste for facilitating another function. By the nature, pretreatment facilities require a higher level of management than that of waste storage facilities.

Anaerobic lagoons are frequently used for pretreatment of agricultural wastes. The lagoon effluent can be treated by various biological, physical, and chemical processes. This chapter introduces anaerobic lagoons, their applications, limitations, performance, reliability, design considerations, energy consumption, capital cost, operation and maintenance cost, and process monitoring notes. Also discussed in this chapter are the waste storage ponds. Sometimes operators want to use lagoon effluent as flush water. To polish and store water for this purpose, waste storage ponds can be constructed in series with the anaerobic lagoon. The capacity of the waste storage pond should be sized for desired storage volume. The application and design procedure is presented in detail.

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2. PROCESS DESCRIPTION

Anaerobic lagoons are relatively deep (up to 20 ft) ponds with steep sidewalls in which anaerobic conditions are maintained by keeping loading so high that complete deoxygenation is prevalent. Although some oxygenation is possible in a shallow surface zone, once grease forms an impervious surface layer, complete anaerobic condition develops. Treatment or stabilization results from thermophilic anaerobic digestion of organic wastes. The treatment process is analogous to that occurring in single stage untreated anaerobic digestion of sludge in which acid forming bacteria break down organics. The resultant acids are then converted to carbon dioxide, methane, cells, and other end products.

In the typical anaerobic lagoon, raw wastewater enters near the bottom of the pond (often at the center) and mixes with the active microbial mass in the sludge blanket, which is usually about 6 ft deep. The discharge is located near one of the sides of the pond, submerges below the liquid surface. Excess undigested grease floats to the top, forming a heat retaining and relatively air tight over. Generally wastewater flow equalization and heating are not practised. Excess sludge is washed out with the effluent. Recirculation of waste sludge is not required. Anaerobic lagoons are capable of providing treatment of high strength wastewaters and are resistant to shock loads. Figure 1 illustrates a typical anaerobic lagoon.

Anaerobic lagoons are customarily contained within earthen dikes. Depending on the soil characteristics, lining with various impervious material such as rubber, plastic, or clay may be necessary. Pond geometry may vary, but surface area to volume ratios is minimized to enhance heat retention.

3. APPLICATIONS AND LIMITATIONS

Although anaerobic biological processes are common for sludge digestion, anaerobic lagoons for wastewater treatment have found only limited applications. The anaerobic lagoon processes are well demonstrated for stabilization of highly concentrated organic wastes, such as animal wastes. Anaerobes lagoons are currently accepted in the United States for the treatment of various animal wastes. Anaerobic treatment of animal waste helps to protect water quality by reducing much of the organic concentration, such as biological oxygen demand (BOD_5), chemical oxygen demand (COD), of the waste. Anaerobic lagoons also reduce the nitrogen content of the waste through ammonia volatilization and effectively reduce animal waste odor if the lagoon is managed properly. Anaerobic lagoons are also effective as treatment units before aerobic treatment of high strength waste. Typically the anaerobic lagoons are used in series with aerobic or facultative lagoons. Anaerobic lagoon process may generate odor. It requires relatively large land area. For efficient operation, water temperature more than 75°F should be maintained.

4. EXPECTED PROCESS PERFORMANCE AND RELIABILITY

BOD_5 removals of 50–70% are achievable depending on loading and temperature conditions. Total suspended solid concentrations may increase, especially if the influent BOD_5 is primarily dissolved. Generally it does not produce an effluent suitable for direct discharge to receiving water. The process is generally resistant to upsets. It is highly reliable if pH in the relatively narrow optimum range is maintained. Anaerobic lagoons may

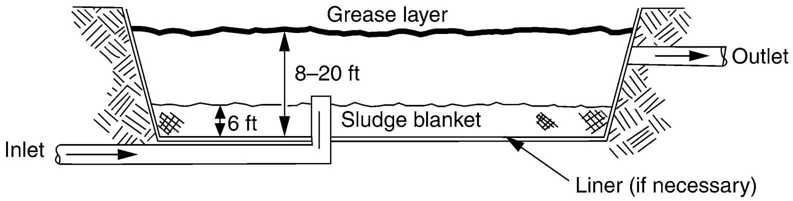


Fig. 1. Cross-section of a typical anaerobic lagoon.

create odor. The lagoons have relatively high land requirements. There is potential for changing wastewater into groundwater unless lagoon is lined. In anaerobic lagoons, excess sludge is usually washed out in the effluent. As anaerobic lagoons are often used for preliminary treatment, recirculation or removal of sludge is generally not required.

5. PROCESS DESIGN

5.1. Minimum Treatment Volume

The maximum operating level of an anaerobic lagoon is a volume requirement plus a depth requirement. The volume requirement is the sum of the following volumes:

a. Minimum treatment volume, MTV (ft^3).
b. Manure volume and wastewater volume WV (ftv^3).
c. Sludge volume, SV (ftv^3).

Polluted runoff from a watershed must not be included in an anaerobic lagoon unless a defensible estimate of the volatile solid loading can be made. Runoff from a watershed such as a feedlot, is not included in an anaerobic lagoon because loading would only result during storm events and because the magnitude of the loading would be difficult, but not impossible, to estimate. As a result, the lagoon would be shocked with an overload of volatile solids.

An automatic outflow device, pipe, or spill way must be placed at a height above the maximum operating level to accommodate the following depths:

a. Normal precipitation less evaporation on lagoon surface (ft).
b. The 25 yr, 24-h storm precipitation on lagoon surface (ft).

These depths added to the depth of the volume requirement of the lagoon establishes the level of the outflow device, pipe, or spillway. The depth of head required to operate the outflow plus a minimum of 1 ft of freeboard is provided above the outflow and establishes the top of the embankment. Should state regulation preclude the use of an outflow device, pipe, or spillway or if for some other reason the lagoon will not have these, storing the 25 yr, 24-h storm precipitation on the anaerobic lagoon surface (a second time) replaces the head requirement. The combinations of these volumes and depths are illustrated in Fig. 2. The terms and derivation are explained in the following paragraphs (1–11).

Anaerobic lagoons are designed on the basis of volatile solids loading rate (VSLR) per 1000 ft^3. Volatile solids represent the amount of solids material in the wastes that will decompose as opposed to the mineral (inert) fraction. The rate of solids decomposition in anaerobic lagoons is a function of temperature; therefore, the acceptable VSLR varies

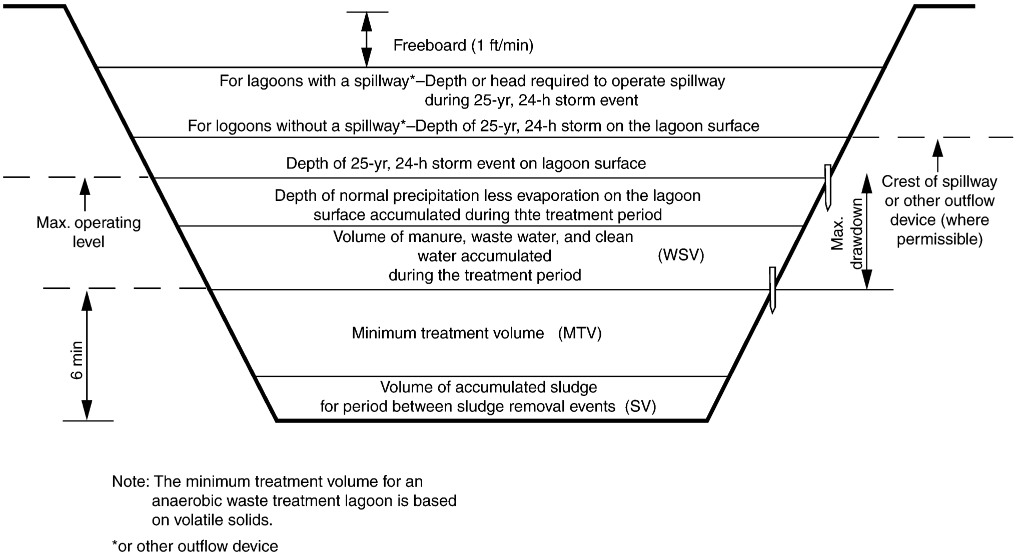


Fig. 2. Illustration of volumes and depth requirements for anaerobic lagoons.

from one location to another. Figure 3 indicates the maximum VSLRs for the United States. If odor need to be minimized, VSLR should be reduced by 25–50%. The minimum treatment volume (MTV) represents the volume needed to maintain sustainable biological activity. The minimum treatment volume for VS can be determined using Eq. (1).

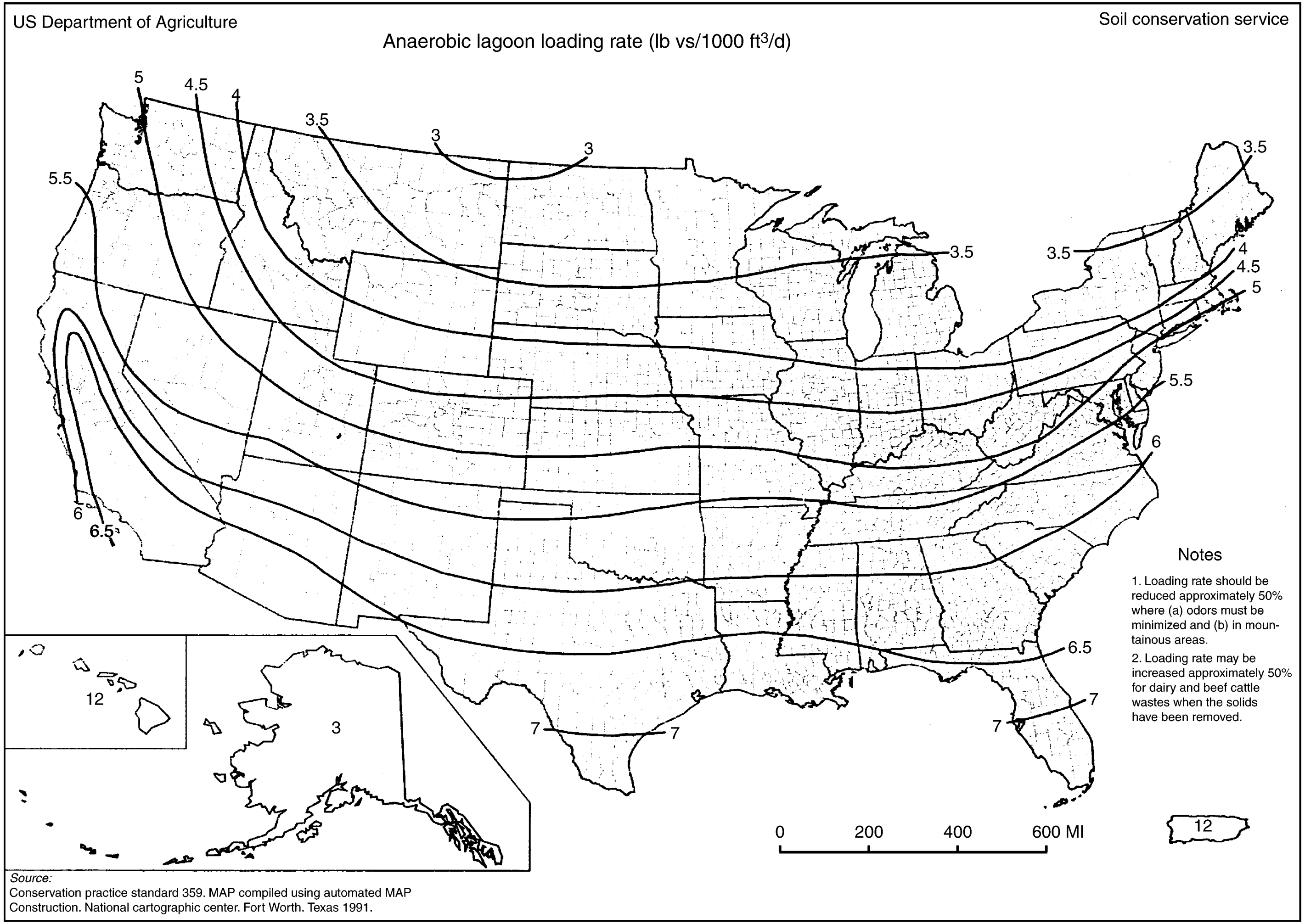
$$\mathrm{MTV} = \frac{\mathrm{TVS}}{\mathrm{VSLR}} \tag{1}$$

where MTV is the minimum treatment volume (ft^3); TVS is the total daily volatile solids loading from all sources (lb/d); and VSLR is the volatile solids loading rate (lb/1000 ft^3/d) (This can be obtained from Fig. 3).

5.2. Waste Volume for Treatment Period

Daily volatile solids production for various wastes can be determined using waste volume for treatment period. If feed spillage exceeds 5%, VSP should be increased by 4% for each additional 1% spillage. Waste volume (WV) should reflect the actual volume of manure, waste water, flash water that will not be recycled, and clean dilution water added to the lagoon during the treatment period. The treatment period is either the detention time required to obtain the desired reduction of pollution potential of the waste or the time between land application events, whichever is longer. State regulations may govern the minimum detention time. Generally, the maximum time between land application events determines the treatment period because this time generally exceeds the detention time required.

$$\mathrm{WV} = \mathrm{TVM} + \mathrm{TWW} + \mathrm{CW} \tag{2}$$



Notes

1. Loading rate should be reduced approximately 50% where (a) odors must be minimized and (b) in mountainous areas.

2. Loading rate may be increased approximately 50% for dairy and beef cattle wastes when the solids have been removed.

Source:
Conservation practice standard 359. MAP compiled using automated MAP Construction. National cartographic center. Fort Worth. Texas 1991.

Fig. 3. Anaerobic lagoons volatile solids loading rates in United States.

Table 1
Sludge Accumulation Ratios

Animal type	Sludge accumulation ratio
Poultry	
Layers	0.0295
Pullets	0.0455
Swine	0.0185
Dairy cattle	0.0729

Source: USDA.

where WV is the waste volume for treatment period (ft^3); TVM is the total volume of manure for treatment period (ft^3); TWW is the total volume of wastewater for treatment period (ft^3); and CW is the clean water added during treatment period (ft^3).

5.3. Sludge Volume

As the manure is decomposed in the anaerobic lagoon only part of the total solids (TS) is reduced. Some of the TS are mineral material that will not decompose, and some of the VS require a long time to decompose. These materials, referred to as sludge, gradually accumulate in the lagoon. To maintain the minimum treatment volume (MTV), the volume of sludge accumulation over the period of time between sludge removals must be considered. Anaerobic lagoons are commonly designed for 15–20-yr sludge accumulation period. The sludge volume (SV) can be determined using Eq. (3).

$$\mathrm{SV} = 365 \times \mathrm{AU} \times \mathrm{TS} \times \mathrm{SAR} \times T \tag{3}$$

where SV is the sludge volume (ft^3); AU is the number of 1000-pound animal units, T is the sludge accumulation time (yr); TS is the total solids production per animal unit per day (lb/AU/d); and SAR is the sludge accumulation ratio (ft^3/lb TS).

5.4. Lagoon Volume Requirement

Total solids values can be obtained from the site investigations. Sludge accumulation ratios should be taken from Table 1. An SAR is not available for beef, but it can be assumed as similar to that for dairy cattle.

The lagoon volume (LV) requirements are for accommodation of the minimum pretreatment volume, the sludge volume, and the waste volume for the treatment period. This is expressed in Eq. (4).

$$\mathrm{LV} = \mathrm{MTV} + \mathrm{SV} + \mathrm{WV} \tag{4}$$

where LV is the lagoon volume requirement (ft^3); MTV is the minimum treatment volume (ft^3) (*see* Eq. [1]); SV is the sludge volume accumulation for period between sludge removal events (ft^3) (*see* Eq. [3]); and WV is the waste volume for the treatment period (ft^3) (*see* Eq. [2]).

In addition to the anaerobic lagoon volume requirement, a provision must be made for depths to accommodate the normal precipitation less evaporation on the anaerobic lagoon surface; the 25-yr, 24-h storm precipitation; the depth required to operate the emergency outflow; and freeboard. Normal precipitation on the lagoon surface is based on the critical treatment period that produces the maximum depth. This depth can be offset to some degree by evaporation losses on the lagoon surface. The offset varies, according to the climate of the region, from a partial amount of the precipitation to an amount in excess of the precipitation. Precipitation and evaporation can be determined from local climate data. Figure 4 shows the average evaporation data in the United States.

The minimum acceptable depth for anaerobic lagoon is 6 ft, but in colder climate at least 10 ft is recommended to assure proper operation and odor control. The design height of an embankment for an anaerobic lagoon should be increased by the amount needed to ensure that the design elevation is maintained after settlement. This increase should not be less than 5% of the design fill height. The minimum top width of the lagoon should be as shown in Table 2.

The combined side slopes of the settled embankment should not be less than 5:1 (horizontal to vertical). The inside slopes can vary from 1:1 for excavated slopes to 3:1 or flatter where embankments are used. Construction technique and soil type must also be considered. In some situation a steep slope may be used below the design liquid level, while a flatter slope is used above the liquid level to facilitate maintenance and bank stabilization. The minimum elevation of the top of the settled embankment should be 1 ft above the maximum design water surface of the lagoon.

5.5. Anaerobic Lagoon Design Criteria

Figure 5 shows a two-lagoon system. Important criteria for designing an anaerobic lagoon system are summarized in below.

a. **Operation.** Parallel or series.
b. **Detention time.** 20–180 d.
c. **Depth.** 8–20 ft.
d. **pH.** 6.8–7.2.
e. **Water temperature range.** 35–120°F.
f. **Optimum waste temperature.** 86°F.
g. **Organic loading.** 200–2200 lb BOD_5/acre/d.
h. **Nutrient requirement.** Nutrient as needed to make up deficiencies in raw wastewater. No other chemical required.
i. **Leakage prevention.** A lagoon should be constructed to avoid leakage and potential in ground water pollution.
j. **Overtopping prevention.** If overtopping can cause embankment failure, an emergency spillway or over flow pipe should be provided. A lagoon can have an over flow to maintain a constant liquid level if the overflow liquid is stored in waste storage pond or otherwise properly managed.
k. **Inlet antifreezing protection.**
l. **Sludge removal.** Sludge removal is an important consideration in the design. This can be accomplished by agitating the lagoon and pumping out the mixed sludge or by using a dragline for removal floating or settled sludge.

5.6. Data Gathering and Compilation for Design

Anaerobic lagoons can be used for treatment of both animal wastes and wastewater. However, the major application of anaerobic lagoons is for animal waste treatment.

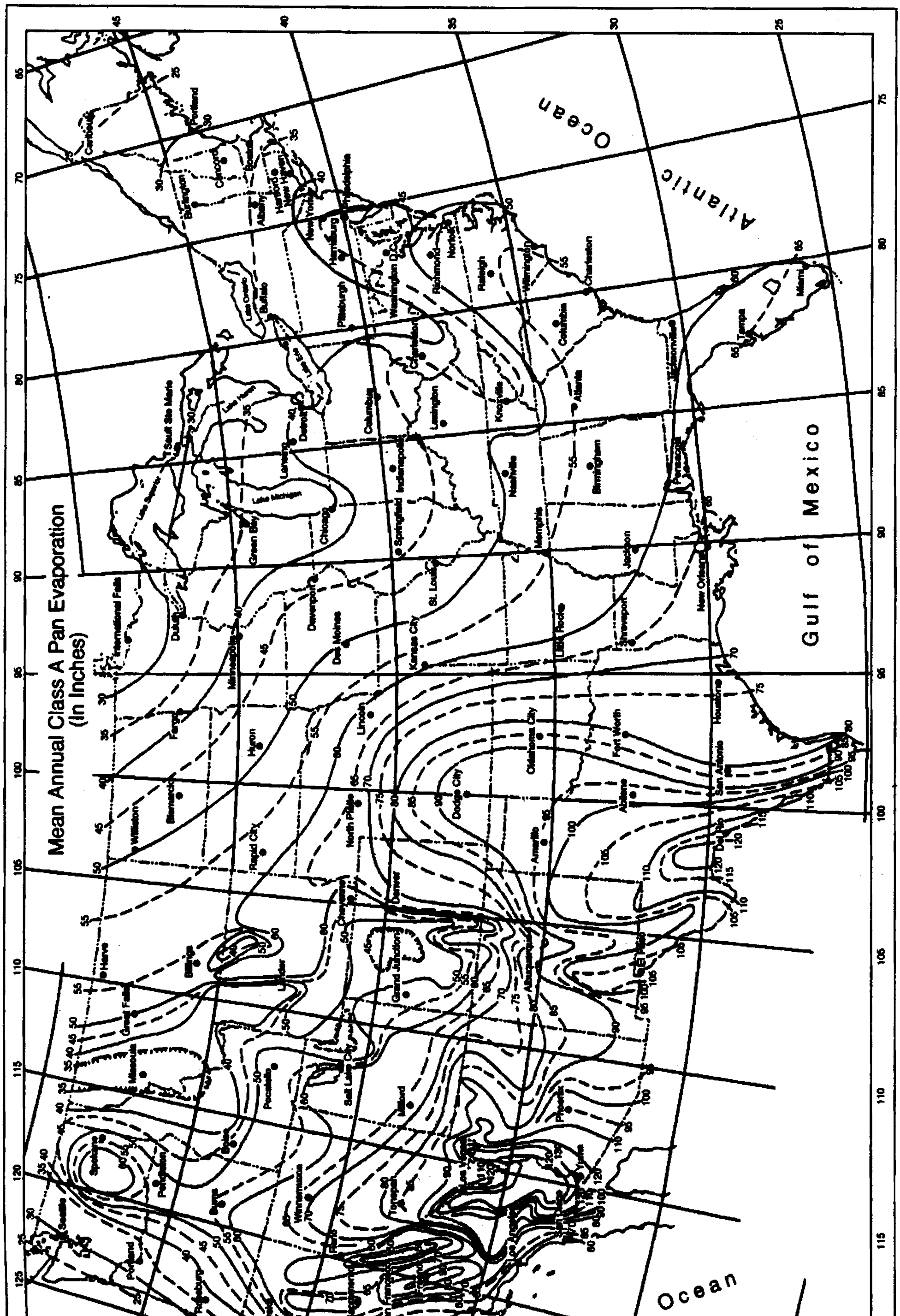


Fig. 4. Average evaporation data in the United States.

Table 2
Minimum Top Width for Lagoon Embankments

Maximum length of embankment (*n*)	Top width (ft)
10 or less	6
11–14	8
15–19	10
20–24	12
25–34	14
34 or more	15

Source: USDA.

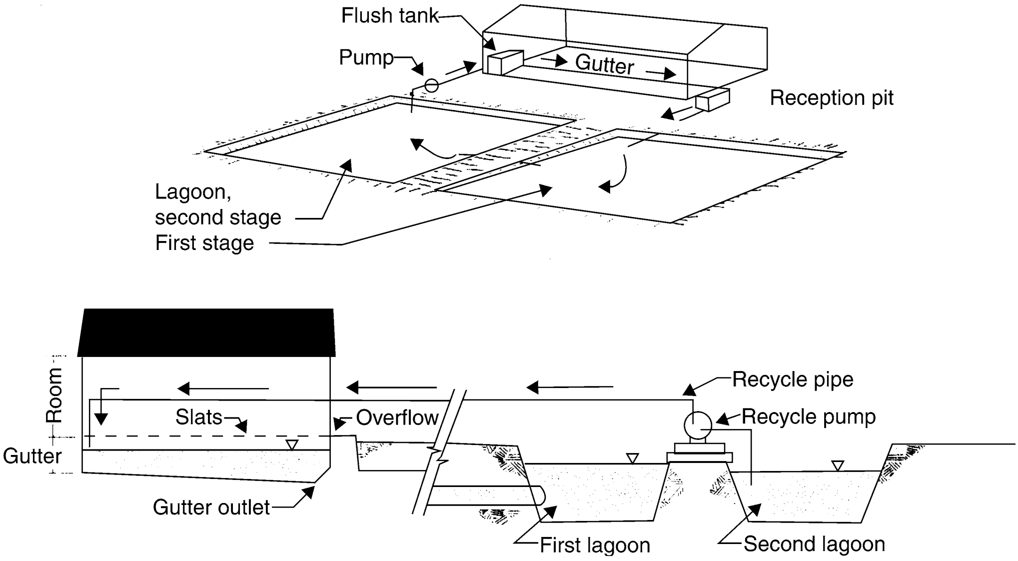


Fig. 5. Two anaerobic lagoons with recycle system.

In case an agricultural waste treatment system is to be developed for an animal farm, the following information should be gathered:

a. Type of animal.
b. Design population of animal.
c. Average weight of each animal (lbs).
d. 25-yr 24-h storm for the local area (in).
e. Net precipitation = precipitation – evaporation (in).
f. Time interval between lagoon pumping = treatment period (d).
g. Time interval between sludge removal (yr).
h. Daily volume of daily manure production (ft^3/AU/d).
i. Daily wastewater volume per animal unit (ft^3/AU/d).
j. Clean water added during treatment period (ft^3).
k. Daily manure to all solids production (lbs/AU/d).
l. Percent volatile content in the total solid manure (%).
m. Lagoon volatile solids loading rate (lb VS/1000 ft^3).
n. Sludge accumulation ratio (ft^3/lb TS).

Table 3
Construction Cost of Anaerobic Lagoons

Wastewater flow (MGD)	Construction cost USD 10^6 (January 2002)
0.1	0.1246
0.5	0.4005
1	0.7120
5	2.4475
10	3.8938
50	13.3500
100	22.2500

Source: US EPA.

Table 4
Operation and Maintenance Cost of Anaerobic Lagoons

Waste water flow (MGD)	Annual OM Costs USD 10^6 (January 2002)
0.1	0.0040
0.5	0.0100
1	0.0156
5	0.0445
10	0.0690
50	0.2069
100	0.3115

Source: US EPA.

o. Sludge accumulation period (yr).
p. Anaerobic lagoon's side slope ratio (horizontal to vertical ratio).

6. ENERGY CONSUMPTION AND COSTS OF ANAEROBIC LAGOONS

Anaerobic lagoons are operated by gravity glow and, therefore have no energy requirement other than any pumping that may be necessary to lift the influent wastewater into the lagoons. Table 3 shows the construction cost (January 2002 USD; ENR CC index = 6390.21) and Table 4 shows the operation and maintenance costs under the following engineering assumptions:

a. January 2002 USD.
b. Service life: 50 yr.
c. Average detention time = 35 d.
d. Depth =10 ft.
e. BOD_5 loading = 466 lb/acre/d.
f. Construction cost includes excavating, grading, and other earthwork and service roads.
g. Costs do not include land and pumping.
h. Operation and maintenance cost consist of labor and material.
i. Waste water characteristics: influent BOD_5, 600 mg/L; effluent BOD_5, 240 mg/L.

To adjust costs for other BOD_5 loading and/or detention times, enter the tabulated data at effective flow.

$$Q = Q_{design} \times \frac{(466\ \text{lb/acre}/d)\ (\text{new detention time})}{(\text{new design loading})\ (35\ d)} \tag{5}$$

It should be noted that the above data of construction cost and OM cost have been complied by the US Environmental Protection Agency (US EPA) for wastewater treatment using anaerobic lagoons. Whether or not the same data can be applied to animal waste treatment remains unknown.

7. WASTE STORAGE PONDS

7.1. Process Description

Sometimes operators want to use lagoon effluent as flush water. To polish and store water for this purpose, waste storage ponds can be constructed in series with the anaerobic lagoon. Storage ponds are earthen basins designed to store wastewater, sludge and manure (Fig. 6). They generally are rectangular, but may be circular or any other shape that is practical for operation and maintenance. The capacity of the waste storage pond should be sized for the desired storage volume. A minimum capacity of the waste storage ponds is the volume for rainfall, runoff, and emergency storm storage. By limiting the depth to less than 6 ft, the pond will function more nearly like an aerobic lagoon. Odor and the level of ammonia, ammonium, and nitrate will be more effectively reduced.

Although, earthen storage is frequently the least expensive type of storage of sludge and manure, it has certain restrictions such as limited space availability, high precipitation, water table, permeable soil, or shallow bedrock, can limit the types of storage considered.

7.2. Process Design

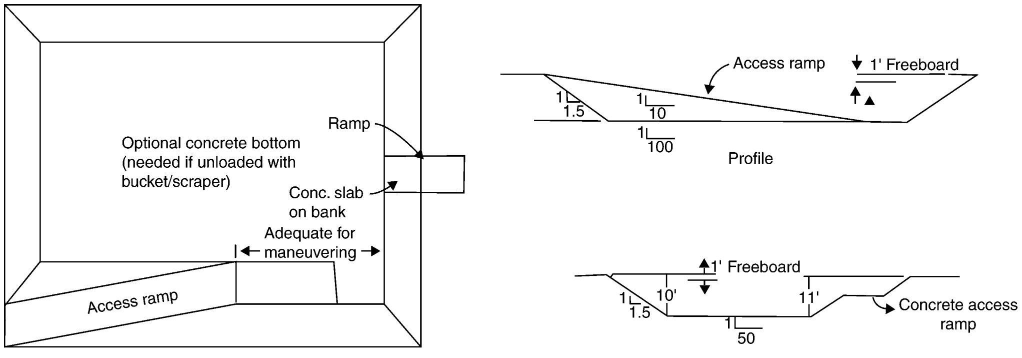
Liquid waste storage ponds and structures should be sized to hold all of the manure, bedding, and wastewater from milkhouse, flushing, and contaminated runoff that can be expected during the storage period. Equation (6) can be used to compute the waste storage volume:

$$\text{WSV} = \text{TVM} + \text{TWW} + \text{TBV} + \text{CW} + \text{ROV} + \text{VSA} \tag{6}$$

where WSV is the waste storage volume for storage period (ft^3); TVM is the total volume of manure for storage period (ft^3); TWW is the total wastewater volume for storage period (ft^3); TBV is the total bedding volume for storage period (ft^3); CW is the clean water added during storage period (ft^3); ROV is the runoff volume (ft^3); and VSA is the solids accumulation volume (ft^3).

Figure 7 shows the cross section of a waste storage pond without a watershed; while Fig. 8 shows the cross section of a waste storage pond with a watershed. Various parameters such as ROV, TVM, CW, TWW, and VSA are clearly illustrated.

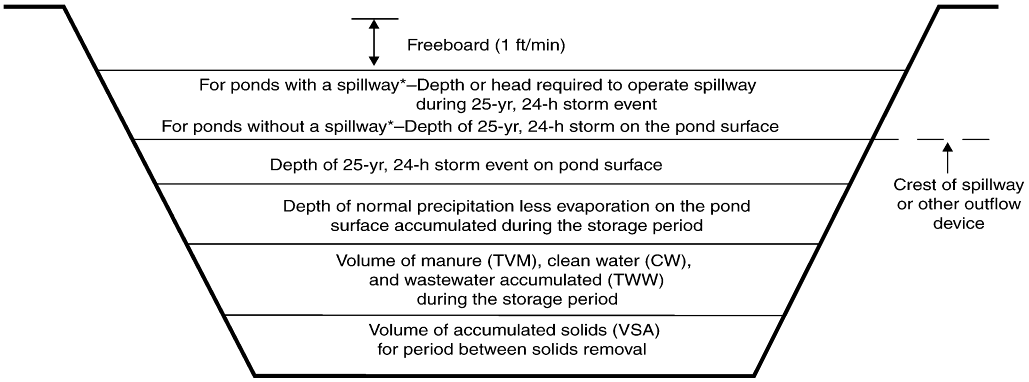
In addition to the waste storage volume, waste storage facilities must, if uncovered, provide a depth to accommodate precipitation less evaporation on the storage surface during the most critical storage period. The most critical storage period is generally the



Plan

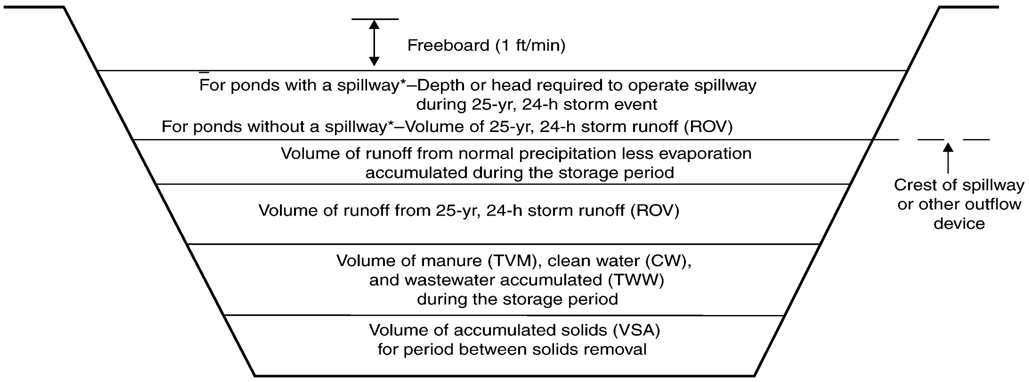
Cross section

Fig. 6. Layout of waste storage pond.



*or other outflow device.

Fig. 7. Cross-section of a waste storage pond without a watershed.



*or other outflow device.

Fig. 8. Cross-section of a waste storage pond with a watershed of a waste storage pond with a watershed.

consecutive months that represent the storage period that gives the greatest depth of the precipitation less evaporation. Frequently, waste storage ponds are designed to include outside runoff from watershed. For these, the runoff volume of the 25-yr, 24-h storm must be included in the storage volume, if the pond does not have a spillway or other outflow device, the runoff volume of 25-yr, 24-h storm must be included a second time.

Accordingly the total depth of a waste storage pond can be estimated as summation of the following:

a. Pond depth calculated based on minimum storage volume (WSV).
b. Added depth resulting from "precipitation less evaporation" for the storage period.
c. Added depth resulting from 25-yr, 24-h storm (only for the ponds without a drainage area).
d. Added depth required to operate emergency outflow.
e. Added depth for freeboard (1 ft minimum).

8. DESIGN AND APPLICATION EXAMPLES

8.1. Example 1

For lagoon sizing and design, how can the volume of a rectangular lagoon be calculated?

Solution

The rectangular lagoon volume can be calculated by the following Eq. (7).

$$V = \frac{\left(4 \times Z^2 \times d^3\right)}{3} + \left(Z \times \mathrm{BL} \times d^2\right) + \left(Z \times \mathrm{BW} \times d^2\right) + \left(\mathrm{BW} \times \mathrm{BL} \times d\right) \tag{7}$$

where V is the lagoon volume (ft^3); Z is the side slope ratio (horizontal to vertical); d is the lagoon depth (ft); BW is the lagoon bottom width (ft); and BL is the lagoon bottom length (ft).

8.2. Example 2

How can the volume of a circular lagoon be calculated?

Solution

The circular lagoon volume can be determined by Eq. (8):

$$V = \left(1.05 \times Z^2 \times d^3\right) + \left(1.57 \times BD \times Z \times d^2\right) + \left(0.79 \times BD^2 \times d\right) \tag{8}$$

where V is the lagoon volume (ft^3); Z is the side slope ratio; d is the lagoon depth (ft); and BD is the lagoon bottom diameter (ft).

8.3. Example 3

Develop a step-by-step design procedure for designing an anaerobic lagoon system to treat the manures from an agricultural farm.

Solution

Step1. Determine animal units.

a. Animal type……………………..
b. Animal weight (W)…………..lb.

c. Number of animal (*N*)…………………………..

d. Animal units (AU) = *W* (*N*)/1000 = …………

Step 2. Determine manure volume.

a. Daily volume of daily manure production per AU (DVM) =………………ft^3/AU/d.

b. Treatment period (*D*) =………………………d.

c. Total volume of manure production for animal type for treatment period VMD = AU × DVM × *D* =………..ft^3.

d. Total manure production for treatment period (TVM) = ……………………ft^3.

Step 3. Determine wastewater volume.

a. Daily wastewater volume per AU (DWW) = ………………………ft^3/AU/d.

b. Total wastewater volume for animal description for treatment period (WWD) = DWW × AU × *D* = …………………….. ft^3.

c. Total wastewater volume for treatment period (TWW) = ………….. ft^3.

Step 4. Determine clean water volume.

a. Clean water added during treatment period (CW) = ……………ft^3.

Step 5. Determine waste volume.

a. Waste volume for treatment period (WV) = TVM + TWW + CW.
= …….. + ……… + ………
= ……………ft^3.

Step 6. Determine the manure.

a. Daily manure total solids production (MTS) = …………………lb/AU/d.

b. Daily manure total solids production for animal type (MTSD).
= MTS × AU = …………..lb/d.

c. Total manure total solids production (TMTS) = …………………..lb/d.

Step 7. Determine manure solids.

a. Daily manure volatile solids production per AU, (MVS) = ………lb/AU/d.

b. Daily manure volatile solids production for animal type per day (MVSD)
= AU × MVS = …………lb/d.

c. Total manure volatile solids production (TMVS) = ……………lb/d.

Step 8. Determine wastewater volatile solids.

a. Daily wastewater volatile solids production (DWVS) = ……………lb/1000 gal.

b. Total wastewater volatile solids production for animal type (WVSD).
= DWVS × DWW × 7.48/(*D* × 1000) = …………………..lb/d.

c. Total wastewater volatile solids production (TWVS) =…………………lb/d.

Step 9. Determine total volatile solids (manure and wastewater).

a. Total daily volatile solids production (TVS) = TMVS + TWVS.
= ……. + ……. = …………lb/d.

Step 10. Determine minimum treatment volume.

a. Select lagoon VS loading rate (VSLR) =………………lb VS/1000 ft^3/d.

b. Minimum treatment volume (MTV) = TVS × 1000/ VSLR
=…….×……. / (…..) =……………ft^3.

Step 11. Determine sludge volume requirement.

a. Sludge accumulation ratio (SAR) = …………ft^3/lb TS.

b. Sludge accumulation period (T) = ………yr.

c. Sludge volume requirement (SV) = 365 × TMTS × T × SAR.
= 365 × …… × … × …… = ………ft^3.

Step 12. Determine minimum lagoon volume requirements.

a. Minimum lagoon volume requirement (MLVR) = MTV + SV + WV.
= ……… + …….. + ………. = ………………ft^3.

Step 13. Determine lagoon size.

a. Side slope ratio (Z) = ………
b. Lagoon volume. Use Eq. (7) or (8) depending on lagoon shape.
c. Lagoon volume (V) must be equal to or more than MLVR = ……………ft^3.
d. Determine the closest lagoon volume.
Trial 1. BW = ……………ft^3; BL = …………ft; d = …………..ft; V = ………….. ft^3.
Trial 2. BW = …….………ft^3; BL = …………..ft; d = …………..ft; V = ………….. ft^3.
Trail 3. BW = ……………ft^3; BL = …………ft; d = …………..ft; V = ………….. ft^3.
Select V = ………………ft^3 ≈ MLVR.

Step 14. Depth adjustment.

a. Depth (d) = ………………ft.
b. Add depth of precipitation less evaporation on lagoon surface for the treatment period = ……………….ft.
c. Add depth of 25-yr, 24-h storm = ……..ft.
d. Add depth required to operate emergency outflow (*Note:* if lagoon design does not include a spillway or other automatic outflow device, use depth of 25-yr, 24-h storm precipitation = ……………ft.
e. Add for freeboard (1 ft minimum) = …………..ft.
f. Final depth = ………………..ft; use ……………ft.

Step 15. Compute total volume of rectangular anaerobic lagoon using final depth.

$$V = \left[\frac{4 \times Z^2 \times d^3}{3}\right] + \left(Z \times \mathrm{BL} \times d^2\right) + \left(Z \times \mathrm{BW} \times d^2\right) + \left(\mathrm{BW} \times \mathrm{BL} \times d\right)$$

=………ft^3.

8.4. Example 4

An animal farm has formally requested assistance in developing an agricultural waste treatment system using an anaerobic lagoon. Assuming you are an environmental engineer, design an anaerobic lagoon system for the animal firm based on the flowing given information:

a. Type of animal = pigs.
b. Design population of animal = 6000.
c. Average weight of animal = 150 lb.
d. 25-yr, 24-h storm for the local area = 6 in.
e. Net precipitation = precipitation – evaporation = 2 in.
f. Time interval between lagoon pumping = treatment period = 180 d.
g. Time interval between sludge removal = 5 yr.
h. Daily volume of daily manure production = 1 ft^3/AU/d.
i. Daily wastewater volume per animal unit = 0 ft^3/AU/d.
j. Clean water added during treatment period = 0 ft^3.
k. Daily manure solids production = 6.34 lb/AU/d.

l. Percent volatile content in the total solids of manure = 85.17%.
m. Lagoon volatile solids loading rate = 6 lb VS/1000 ft^3.
n. Sludge accumulation ratio = 0.0485 ft^3/lb TS.
o. Sludge accumulation period = 5 yr.
p. Anaerobic lagoon's side slope ratio (horizontal to vertical ratio) = 2.

Solution

Because 85.11% of the total solids in the given information item (*R*) is volatile, the daily manure volatile solids production per AU, or MVS, is estimated to be 5.4 lbs/AU/d. The step-by-step design procedures are used for the detailed design as follows:

Step 1. Determine animal units.
a. Animal type is pigs.
b. Animal weight (*W*) 150 lb.
c. Number of animals (*N*) 6000.
d. Animal units (AU) = W (N)/1000 = 900.

Step 2. Determine manure volume.
a. Daily volume of daily manure production per AU (DVM) = 1 ft^3/AU/d.
b. Treatment period (D) = 180 d.
c. Total volume of manure production for animal type for treatment period.
   VMD = AU×DVM × D = 162,000 ft^3.
d. Total manure production for treatment period (TVM) = 162,000 ft^3.

Step 3. Determine wastewater volumes.
a. Daily wastewater volume per AU (DWW) = 0 ft^3/AU/d.
b. Total wastewater volume for animal description for treatment period (WWD).
   = DWW × AU × D = 0 ft^3.
c. Total wastewater volume treatment period (TWW) = 0 ft^3.

Step 4. Determine clean water volume.
a. Clean water added during treatment period (CW) = 0 ft^3.

Step 5. Determine waste volume.
a. Waste volume for treatment period (WV) = TVM + TWW + CW = 162,000 + 0 + 0 = 162,000 ft^3.

Step 6. Determine the manure.
a. Daily manure total solids production (MTS) = 6.34 lb/AU/d.
b. Daily manure total solids production for animal type (MSTD) = MTS × AU.
   = 5706 lb/d.
c. Total manure total solids production (TMTS) = 5706 lb/d.

Step 7. Determine manure solids.
a. Daily manure volatile solids production per AU (MVS) = 5.4 lb/AU/d.
b. Daily manure volatile solids production for animal type per day (MVSD).
   = AU × MVS = 4860 lb/d.
c. Daily manure volatile solids production = (TMVS) = 4860 lb/d.

Step 8. Determine waste water volatile solids.
a. Daily waste water volatile solids production (DWVS) = 0 lb/1000 gal.
b. Total wastewater volatile solids production for animal type (WVSD) = DWVS × DWW × 7.48/(D × 1000) = 0 lb/d.
c. Total wastewater volatile solids production (TWVS) = 0 lb/d.

Step 9. Determine total volatile solids (manure and wastewater).

a. Total daily volatile solids production (TVS) = TMVS + TWVS
   = 4860 + 0 = 4860 lb/d.

Step 10. Determine minimum treatment volume.

a. Select lagoon VS loading rate (VSLR) = 6 lb VS/1000 ft^3/d.
b. Minimum treatment volume (MTV) = TVS × 1000/VSLR = 4860 × 1000/6.
   = 810,000 ft^3.

Step 11. Determine sludge volume requirement.

a. Sludge accumulation ratio (SAR) = 0.0485 ft^3/lb TS.
b. Sludge accumulation period (T) = 5 yr.
c. Sludge volume requirement (SV) = 365 × TMTS × T × SAR.
   = 365 × 5706 × 5 × 0.485 = 505,052 ft^3.

Step 12. Determine minimum lagoon volume requirement.

a. Minimum lagoon volume requirement (MLVR) = MTV + SV + WV.
   = 810,000 + 505,052 + 162,000 = 1,477,052 ft^3.

Step 13. Determine lagoon size.

a. Side slope ratio (Z) = 2.
b. Lagoon volume (V).
   $= [(4 \times Z^2 \times d^3)/3] + (Z \times BL \times d^2) + (Z \times BW \times d^2) + (BW \times BL \times d)$.
c. Lagoon volume (V) must be equal to or more than MLVR = 1,477,052 ft^3.
d. Determine the closest lagoon volume.
   Trail 1. BW = 150 ft; BL = 1000 ft; d = 8 ft; V = 1,344,931 ft^3.
   Trail 2. BW = 150 ft; BL = 1200 ft; d = 8 ft; V = 1,615,531 ft^3.
   Trial 3. BW = 150 ft; BL = 1100 ft; d = 8 ft; V = 1,482,731 ft^3.
   Select V = 1,482,731 ft^3 < MLVR.

Step 14. Depth adjustment.

a. Depth (d) = 8 ft.
b. Add depth of precipitation less evaporation on lagoon surface for the treatment period = 0.6 ft.
c. Add depth of 25-yr, 24-h storm = 0.5 ft.
d. Add depth required to operate emergency outflow (*Note:* If lagoon design does not include a spillway or other automatic outflow device, use depth of 25-yr, 24-h storm precipitation).
   = 0.3 ft.
e. Add for freeboard (1ft minimum) = 1 ft.
f. Final depth = 10.4 ft; use 10.5 ft.

Step 15. Compute total volume of rectangular anaerobic lagoon using final depth.

$(V) = [(4 \times Z^2 \times d^3)/3] + (Z \times BL \times d^2) + (Z \times BW \times d^2) + (BW \times BL \times d)$
   = 2,014,300 ft^3.

8.5. Example 5

An animal farm has 500 Milkers animals each weighing 1400 lb, 150 Dry animal each weighing 1400 lb and 150 Heifers animals each weighing 1000 lb. The daily volume of manure production rates (DVM) are:

a. Milkers = 1.3 ft^3/AU/d.
b. Dry = 1.1 ft^3/AU/d.
c. Heifers = 1.3 ft^3/AU/d.

Determine (a) the total animal units (AU), and (b) the total manure production for intended storage period of 180 d.

Solution

1. *Determination of total animal units.*

$$\mathrm{AU} = W_1\,N_1/1000 + W_2\,N_2/1000 + W_3\,N_3/1000$$
$$= (1400 \times 500)/1000 + (1400 \times 150)/1000 + (1000 \times 150)/1000$$
$$= 700 + 210 + 150 = \mathrm{AU}_1 + \mathrm{AU}_2 + \mathrm{AU}_3.$$

2. *Determination of TVM.*

$$\mathrm{TVM} = \mathrm{AU}_1 \times \mathrm{DVM}_1 \times D + \mathrm{AU}_2 \times \mathrm{DVM}_2 \times D + \mathrm{AU}_3 \times \mathrm{DVM}_3 \times D.$$
$$= 700 \times 1.3 \times 180 + 210{\times}1.1 \times 180 + 150 \times 1.3 \times 180$$
$$= 163800 + 41580 + 35100.$$
$$= 240{,}480\ \mathrm{ft}^3.$$

8.6. Example 6

Determine the waste storage volume for designing a waste storage pond for the same animal farm described in Example 5. Assuming the following is known:

a. Daily wastewater volume per AU for Milkers = 0.6 ft^3/AU/*d*.
b. Daily wastewater volume per AU for Dry = 0.
c. Daily wastewater volume per AU for Heifers = 0.

The clean water volume (CW), runoff volume (ROV) and solids accumulation volume (VSA) are all zero. Total manure production (TVM) is 240,480 ft^3 from Example 5.

Solution

Total wastewater volume for the storage period (TWW) can be determined as follows:

$$\mathrm{TWW} = \mathrm{WWD}_1 + \mathrm{WWD}_2 + \mathrm{WWD}_3.$$
$$= \mathrm{DWW}_1 \times \mathrm{AU}_1 \times D + \mathrm{DWW}_2 \times \mathrm{AU}_2 \times D + \mathrm{DWW}_3 \times \mathrm{AU}_3 \times D.$$
$$= 0.6 \times 700 \times 180 + 0 \times 210 \times 180 + 0 \times 150 \times 180 = 75{,}600\ \mathrm{ft}^3.$$

Waste storage volume (WSV) can then be calculated using Eq. (6).

$$\mathrm{WSV} = \mathrm{TVM} + \mathrm{TWW} + \mathrm{TBV} + \mathrm{CW} + \mathrm{ROV} + \mathrm{VSA}.$$
$$= 240{,}480 + 75{,}600 + 0 + 0 + 0 + 0 = 316{,}080\ \mathrm{ft}^3.$$

8.7. Example 7

Based on the technical information from Examples 5 and 6, design a rectangular waste storage pond, assuming:

a. Waste storage volume (WSV) = 316,080 ft^3.
b. Side slope ratio (Z) = 3.
c. Depth of precipitation less evaporation for the storage period = 2.3 ft.
d. 25-yr, 24-h storm (for ponds without a drainage area) = 0.3 ft.
e. Depth requirement to operate emergency outflow = 0.3 ft.
f. Freeboard requirement = 1 ft min.

Solution

Equation (7) in Example 1 is used for calculating pond volume (V).

Trial 1. BW = 100 ft; BL = 500 ft; d = 6 ft; V = 367,392 ft^3.
Trial 2. BW = 100 ft; BL = 400 ft; d = 6 ft; V = 296,592 ft^3.
Trial 3. BW = 100 ft; BL = 425 ft; d = 6 ft; V = 314,292 ft^3.
Trial 4. BW = 100 ft; BL = 428 ft; d = 6 ft; V = 320,580 ft^3.
Very close to WSV of 316,080 ft^3.

Finally the waste storage pond's depth must be adjusted in order to determine the final depth:

Final depth = 6.1 + 2.3 + 0.3 + 0.3 + 1 = 10 ft.

NOMENCLATURE

AU	Number of 1000-pound animal units
BD	Lagoon bottom diameter (ft)
BL	Lagoon bottom length
BW	Lagoon bottom width
CW	Clean water added during treatment period (ft^3)
d	Lagoon depth (ft)
D	Treatment period (d)
LV	Lagoon volume requirement (ft^3)
MTV	Minimum treatment volume (ft^3), *see* Eq. (1)
N	Number of animals
ROV	Runoff volume (ft^3)
SAR	Sludge accumulation ratio (ft^3/lb TS)
SV	Sludge volume (ft^3)
T	Sludge accumulation time (yr)
TBV	Total bedding volume for storage period (ft^3)
TS	Total solids production per animal unit per day (lb/AU/d)
TVM	Total volume of manure for treatment period (ft^3)
TVS	Total daily volatile solids loading from all sources (lb/d)
TWW	Total wastewater volume (ft^3)
V	Lagoon volume (ft^3)
VSA	Solids accumulation volume (ft^3)
VSLR	Volatile solids loading rate (lb/1000 ft^3/d)
W	Average weight of each animal (lb)
WSV	Waste storage volume for storage periods (ft^3)
WV	Waste volume for treatment period (ft^3)
Z	Side slope ratio (horizontal to vertical)

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Vertical Shaft Digestion, Flotation, and Biofiltration

**Lawrence K. Wang, Nazih K. Shammas, Jeffrey Guild,
and David Pollock**

CONTENTS

1. INTRODUCTION

1.1. *Biosolids Treatment*

Solids processing represents about 40% of the overall costs at a wastewater treatment plant. Biosolids processing refers to the screening, grit removal, thickening, stabilization, dewatering, drying, and disinfection of sludge, and also air emission control, if the target waste contains toxic volatile organic compounds (VOCs), and odorous substances.

This particular chapter introduces three new biosolids treatment processes, vertical shaft digestion (VSD), vertical shaft flotation (VSF) thickening, and gas-phase biofiltration. The combination of these three processes, and a few supplemental ones (such as grit removal, dewatering and drying) provide complete biosolids treatment. Specifically, this chapter discusses biosolids treatment objectives, theory and principles, description of processes vertical shaft bioreactor (VSB), VSD, aerobic digestion, autothermal thermophilic aerobic digestion, VSF thickening, optional anaerobic digestion, biosolids dewatering, and air emission control by biofiltration, engineering design, and case studies.

1.2. *VSB and VSD*

VSB is one of the advanced activated sludge processes for wastewater treatment (1,2), while VSD is one of advanced aerobic digestion processes for biosolids treatment (3,4). Both VSB and VSD are alike from structural view points. Similarly, activated sludge process and aerobic digestion are similar to each other in terms of physical structure. Both VSB and VSD involve the use of vertical shaft reactors, which are typically

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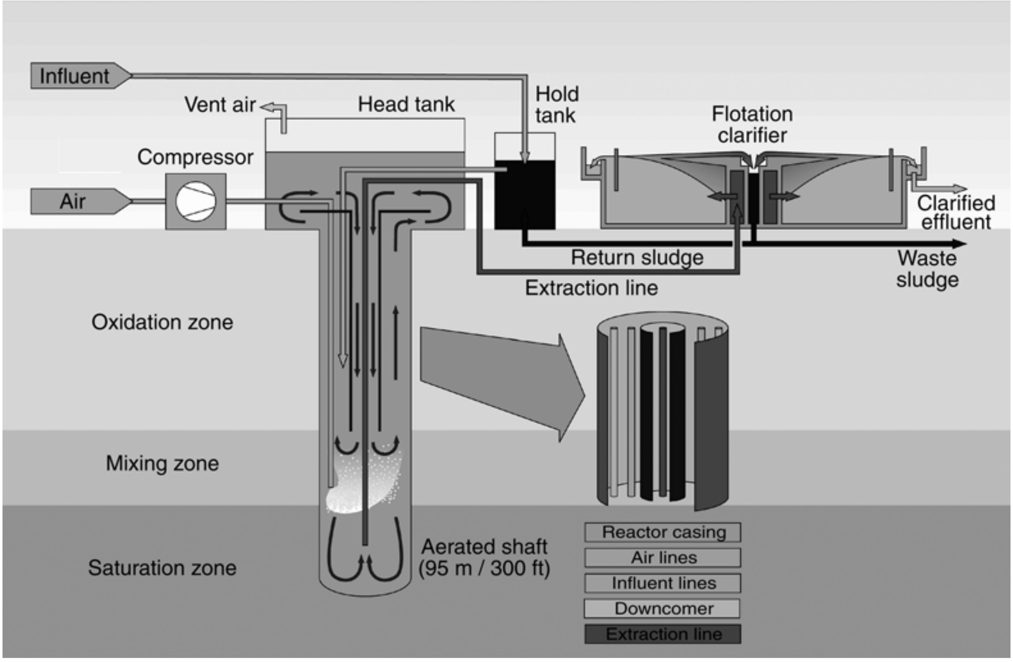


Fig. 1. Vertical shaft bioreactor (VERTREAT™) process flow diagram.

350–500 ft in depth, and 2.5–10 ft in diameter. Although vertical shaft reactors are usually constructed in ground, they can be constructed above ground when necessary. Figures 1 and 2 show a VSB system for wastewater treatment and a VSD system for biosolids treatment, respectively.

When the vertical shaft reactor is used in a VSB system for wastewater treatment (Fig. 1), the reactor is mainly an aeration unit, and its influent is usually the primary wastewater effluent. The VSB effluent is treated by either a sedimentation or flotation clarifier, although the later is preferred. The clarifier effluent is then subjected to additional disinfection treatment, or advanced treatment (such as tertiary granular activated carbon adsorption, tertiary filtration, ion exchange, ultraviolet oxidation/disinfection, and so on) before its discharge to receiving water. Typical commercial VSB processes include VERTREAT™ and DEEP SHAFT, both of which have been extensively used in the UK, US, and Japan. The readers are referred to other sources for detailed information on VSB processes (1,2,5–9).

When the vertical shaft reactor is used in a VSD system for biosolids treatment (Fig. 2), it becomes a vertical shaft autothermal thermophilic aerobic digester (VSD–ATAD) and its influent is usually a thickened sludge stream. A typical commercial VSD process is the VERTAD™ system manufactured by NORAM Engineering and Constructors Ltd., Vancouver, Canada. A VSD system for biosolids treatment can be operated as a sequencing batch reactor (10,11), or as a continuous biological digestion process. Either air or pure oxygen can be used for biosolids digestion/oxidation; therefore, VSD can be either a VSD–autothermal thermophilic aerobic digestion using air (ATAD–air), or a VSD–autothermal thermophilic aerobic digestion using oxygen (ATAD–oxygen) process. The VSD effluent is usually discharged to a flotation thickening unit and a dewatering

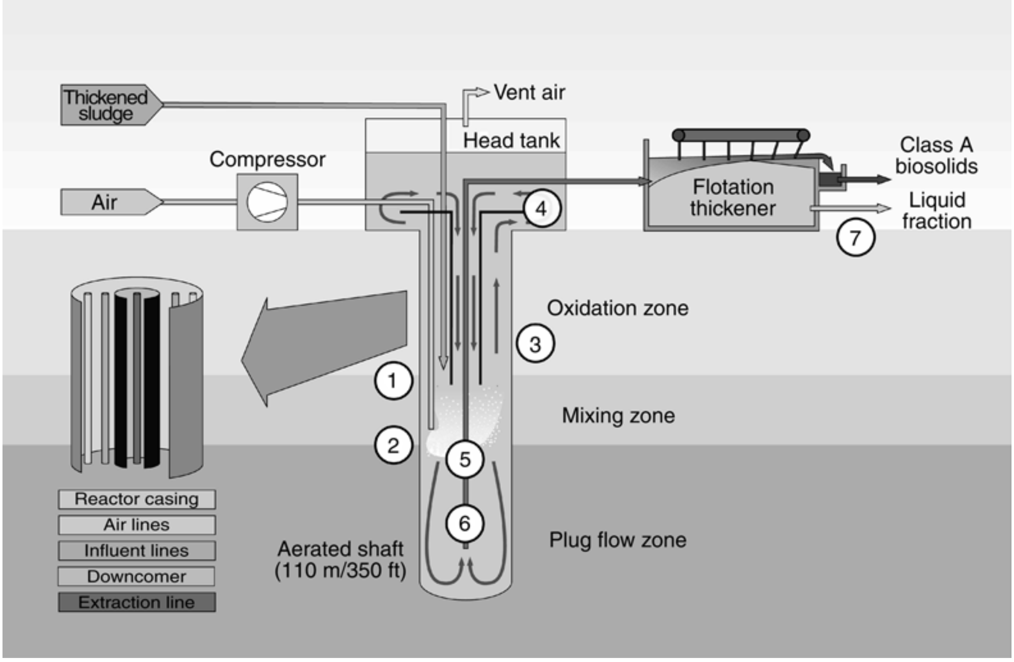


Fig. 2. VSD (VERTAD™) process flow diagram.

unit for further biosolids-water separation. The dewatered biosolids are either reused by spreading on agricultural land or sent to final land disposal (sanitary landfill) or for incineration.

It has been proven that the combined application of aerobic and anaerobic digestion will have a significant improvement upon the biosolids treatment efficiency. Accordingly, VSD may be used for retrofitting the existing mesophilic anaerobic digesters. For this particular reason, the anaerobic digestion process and the results of the combined VSD-anaerobic digestion of biosolids are both briefly introduced in this chapter.

VSD is an advanced aerobic digestion process especially feasible for biosolids treatment in cold climate, where conventional aerobic digestion, or innovative cryophilic aerobic digestion (8,9) are not cost-effective because of their big foot-print and above ground environment. VSD has extremely small foot-print, thus has small heat-loss. In addition, its vertical deep shaft reactor will allow the use of ground heat, and will exhibit a high oxygen transfer efficiency as a result of the high hydraulic pressure (350–500 ft water column). VSD is also an attractive choice for biosolids digestion when the plant runs out of space for future expansion.

1.3. Vertical Shaft Flotation (VSF) Thickening Process

Dissolved air flotation (DAF) is one of the best biosolids thickening processes (12). It is a well-established technology and has widespread applications in wastewater treatment plants. Its only drawback is the high power cost for waste stream pressurization, gas injection and dissolution, gas release, and micro-gas bubble generation. VSF thickening process is always used in conjunction with the VSD. The deep bioreactor serves as a pressurization tank for both gas injection and gas dissolution under high hydraulic pressure. The liquid biosolids stream in the bioreactor is supersaturated with dissolved gas. After the VSD effluent is discharged to a flotation tank under normal 1 atm pressure, the extremely fine gas bubbles will be generated because of sudden pressure release. The fine gas bubbles float the biosolids to the water surface forming a scum layer which is then skimmed off for further treatment (such as dewatering). The subnatant is recycled to the plant influent for liquid treatment.

The adoption of VSF saves:

a. Capital costs for pressure tanks and gas injection.
b. Operation and maintenance costs for gas bubble generation.
c. Treatment costs caused by sludge bulking problems.

1.4. Gas-Phase Biofiltration

A liquid-phase biofiltration also known as trickling filter treatment has been used extensively for decades for wastewater treatment (13). It was not until the last decade that biofiltration was modified for treatment of air emission streams aiming at the reduction of biodegradable VOCs and odor-causing substances (14,15).

Gas-phase biofiltration process equipment is commercially available (15). While the capital and operation and maintanence costs of gas-phase biofiltration are affordable, the gas stream collection cost is usually high. For this reason, conventional wastewater and biosolids treatment facilities usually do not have air emission collection and treatment systems installed. A VSD unit has a very small foot print (2.5–10 ft in diameter) making it feasible to install a complete air emission collection and biofiltration system for total environmental control. When wastewater or biosolids are known to contain toxic biodegradable VOCs, one will seriously consider the inclusion of such equipment, a VSB for wastewater treatment, and/or a VSD for biosolids stabilization. Both systems can be totally covered for cost-effective air emission control using biofiltration.

1.5. Biosolids Digestion and Stabilization

Several techniques can be used for biosolids stabilization such as: anaerobic digestion, aerobic digestion, alkaline treatment, and composting. The primary purpose of stabilization is to reduce the biological activity of organic matter in the raw biosolids. Active organic matter can attract disease-carrying vectors such as flies. The secondary goals of stabilization are to reduce the mass of organic solids and the concentration of pathogenic bacteria. Given the multifaceted challenges that managers face, many are looking to advanced digestion to achieve their objectives related to:

a. **Biosolids quality.** Reduce the pathogenic organisms and fecal coliform density.
b. **Solids reduction.** Reduce the amount of residual biosolids requiring hauling or tipping fees.
c. **Digester capacity.** Reduce the volume required for biosolids stabilization.
d. **Life cycle cost.** Reduce the life cycle cost of constructing, operating, and maintaining the digestion facility.
e. **Energy management.** Reduce the plant energy requirements.
f. **Operating characteristics.** Reduce odor, foaming, cleaning frequency, and impacts of side streams on wastewater treatment; improve mixing, heating, gas production for anaerobic digestion, and dewaterability.

To achieve some of these objectives, we can look to high performance or advanced digestion processes. This chapter concentrates on VSD and related flotation thickening, dewatering, and optional anaerobic digestion processes. The key objectives relate to biosolids quality, solids reduction, digester capacity, life cycle cost, energy management, and various operating characteristics. The following subsections are based on the report of the Advanced Digestion Technology Team, Bioenergy Subcommittee, Residuals and Biosolids Committee of Water Environment Federation in 2002 (9).

1.5.1. Biosolids Quality

One of the driving forces for the development of advanced digestion technologies is the increasingly stringent regulatory requirements for biosolids handling. With increasing environmental regulation, rising disposal costs, and more emphasis on public perception, the production of high quality biosolids is becoming more important. High quality biosolids have an improved public perception and a tremendous opportunity for beneficial use through land application (9).

High quality biosolids, as defined by the US Environmental Protection Agency (US EPA) under 40CFR503, are divided into two classifications, class B and class A, based on the level of pathogen reduction achieved by the treatment process. All biosolids that are to be land applied for beneficial use must meet the requirements of one of these classifications. Class B biosolids are usually achieved through a process to significantly reduce pathogens, as defined by 40CFR503 Appendix B (Appendix 1 of this chapter). These processes include aerobic and anaerobic digestion, composting, lime stabilization, composting, and air drying. There are a number of restrictions on the harvesting of food crops, grazing of animals, and public access to land where class B biosolids have been applied (9).

The production of higher quality, class A biosolids offers the advantage of increased flexibility as there are few restrictions on the beneficial use or sale of class A biosolids. In order to produce class A biosolids, one of the seven alternatives listed in 40CFR503.32 must be met. The production of class A biosolids through advanced digestion usually falls under Alternatives 1, 5, and 6.

To meet the requirements of Alternative 1, sewage sludge is held at elevated temperatures for a specified amount of time, as determined by equations given in 40CFR503.32. To meet the requirement of Alternative 5, sewage sludge is treated in a process to further reduce pathogens (PFRP) as defined by 40CFR503 Appendix B. Digestion coupled with pasteurization and thermophilic aerobic digestion are defined as PFRP's. Alternative 6 allows the regulating community to determine a given process to be equivalent to a PFRP (9).

1.5.2. Solids Reduction

Biosolids' solids reduction is one of the main objectives for sludge stabilization.

1.5.3. Digester Capacity

Some wastewater treatment plants have limited space available, but need higher digester capacity. Some advanced digestion technologies increase the capacity of existing tanks because of their shorter retention time and small foot print.

1.5.4. Life Cycle Cost

Whether an existing digestion facility is being upgraded or a new facility is being designed, minimizing life cycle cost is an objective.

1.5.5. Energy Management

Energy management involves demand-side management and resource management. Anaerobic digesters are being perceived more as an energy source than just a sludge stabilization technology.

1.5.6. Operating Characteristics

Operating characteristics include odor control, foam control, frequency of cleaning (surface and sediment). The complexity of a digester technology is an issue (9).

2. PRINCIPLES OF VERTICAL SHAFT DIGESTION (VSD) AND OPTIONAL ANAEROBIC DIGESTION

2.1. Theory and Principles of Aerobic Digestion

Biological biosolids digestion is a method of sludge stabilization that uses bacteria to degrade organic matter. The principal purposes of stabilization are to make the treated biosolids less odorous, and to reduce the pathogenic organism content. Digestion also results in a substantial decrease in the mass of suspended biosolids.

There are seven different kinds of biological aerobic digestion processes (3,4,9,16–29), which are as follows:

a. Conventional aerobic digestion using air (AD-air).
b. Conventional aerobic digestion using oxygen (AD-oxygen).
c. Conventional autothermal thermophilic aerobic digestion using air (ATAD-air).
d. Conventional autothermal thermophilic aerobic digestion using oxygen (ATAD-oxygen).
e. VSD using air (VSD–ATAD-air).
f. VSD using oxygen (VSD–ATAD-oxygen).
g. Cryophilic aerobic digestion.

VSD is an ATAD process using either air or oxygen; therefore, VSD can be either VSD–ATAD-air or VSD–ATAD-oxygen process. Both VSD–ATAD-air and VSD–ATAD-oxygen use vertical shaft reactors (350–500 ft in depth), whereas both ATAD-air and ATAD-oxygen use conventional shallow bioreactors. The theory and principles of all aerobic digestion processes are alike and will be briefly covered in this section. For further details the readers are referred to another chapter (4).

The biological aerobic digestion process involves the direct oxidation of biodegradable matter and microbial cellular material by a biologically active mass of organisms. This is illustrated by the following reactions in the presence of microorganisms (13,30–37):

$$\text{Organic matter} + O_2 \rightarrow \text{cellular matter} + CO_2 + H_2O \tag{1}$$

$$\text{Cellular matter} + O_2 \rightarrow \text{digested sludge} + CO_2 + H_2O \tag{2}$$

Normally, the second reaction (called endogenous respiration) is the predominant reaction in aerobic digestion (30–37). Endogenous respiration is the process whereby microorganisms metabolize their own protoplasm without replacement. Stabilization is not complete until there has been an extended period of primarily endogenous respiration (15–20 d for conventional aerobic digestion, and 4–6 d for VSD), if conventional aerobic digestion is applied. VSD is operated under high pressure enhancing oxygen transfer and bio-oxidation, thus significantly reducing the required retention time for biosolids stabilization. Although the biochemical reactions shown in Eqs. (1) and (2) hold true for all aerobic digestion processes, the true mechanisms and kinetics of pressurized biochemical reactors, such as vertical shaft reactors are still unknown.

It is important to note that the nitrification (36) occurs in the mesophilic aerobic digestion processes (such as AD-air and AD-oxygen), but does not occur in the autothermal thermophilic aerobic digestion processes (such as, ATAD-air, ATAD-oxygen, VSD–ATAD-air, and VSD–ATAD-oxygen). Small-scale aerobic digestion systems often use a one-tank sequencing batch system (11) with a complete mix cycle followed by settling and decanting (to help thicken the sludge). Larger operations may use a separate sedimentation tank to allow continuous flow and facilitate decanting and thickening. Either air or pure oxygen can be used in these systems. The aerobic digestion process is less sensitive to environmental factors than anaerobic digestion. However, one of its limitations is that it has less established design parameters.

2.2. *Theory and Principles of Optional Anaerobic Digestion*

VSD may be used to retrofit an existing failed mesophilic anaerobic digestion to form a combined system; therefore, the theory and principles of anaerobic digestion are also briefly covered here (2,4,38–40). Anaerobic digestion can be a single stage or a two stage digestion system. It can be operated as either a mesophilic anaerobic digestion or a thermophilic anaerobic digestion. The basic theory and principles of all anaerobic digestion processes are alike. Briefly speaking, anaerobic digestion is performed by several groups of anaerobic and facultative organisms that simultaneously assimilate and break down organic matter. It is a two-phase process. First, acid-forming organisms convert the organic substrate to volatile organic acids. Little change occurs in the total amount of organic material in the system, although some lowering of pH results. Alkaline buffering materials are also produced. Next, the volatile organic acids are converted primarily to methane and carbon dioxide.

This anaerobic process is essentially controlled by the methane-producing bacteria. These bacteria grow at a relatively slow rate and have generation times which range from slightly less than 2 d to about 22 d. Methane formers are very sensitive to pH, substrate composition, and temperature. If the pH drops less than 6.0, methane formation ceases, and there is no decrease in organic content of the sludge. The methane bacteria are highly active in the mesophilic and thermophilic ranges. The mesophilic range is between 10 and 47°C (50 and 110°F), whereas the thermophilic range is between 45 and 65°C (113 and 149°F). Essentially, almost all digesters in the United States operate within the mesophilic temperature range.

Although very widely used, anaerobic digesters are sensitive to a variety of physical, chemical, and biological phenomena (e.g., pH, alkalinity, temperature, and concentrations of toxic substances). Anaerobic sludge digester biomass is relatively intolerant to changing environmental conditions. The process requires careful monitoring of pH, gas production, and volatile acids.

Anaerobic digestion can be performed in one or two stages. In single stage systems one tank is used for digestion and thickening. As decomposition proceeds, three distinct zones develop: the scum layer at the top of the digester, the supernatant zone in the middle and the sludge zone at the bottom. The sludge zones include an actively decomposing upper layer and a relatively stabilized bottom layer where the stabilized sludge accumulates. Two-stage anaerobic digestion evolved as an attempt to provide additional gas production as well as a separate settling and thickening process in the secondary digester. The readers are referred to another book of this handbook series for more detailed technical information on anaerobic digestion (41).

2.3. Combined VSD and Anaerobic Digestion

Digestion reduces sludge volumes and produces less odorous biosolids that are often easier to dewater. VSD is an aerobic digestion process which has some advantages over anaerobic digestion including simplicity of operation, lower capital cost, fewer effects from interfering substances (such as heavy metals), and no danger of methane explosions. Because anaerobic digestion has a higher ability of reducing volatile solids content than aerobic processes and as it has the advantage of producing methane as an energy source, VSD may be applied in conjunction with anaerobic digestion (9).

Pilot testing is recommended before using a combined aerobic-anaerobic biosolids digestion system to confirm and/or select the design and operating parameters. The primary result of combined aerobic and anaerobic digestion system is the efficient reduction of volatile solids. The performance of aerobic digestion (such as VSD) depends on detention time, temperature, and character of solids. The performance of anaerobic digestion depends on proper seeding, pH, character of solids, temperature, and degree of mixing of raw solids with actively digesting seed material.

3. DESCRIPTION, OPERATION, AND APPLICATIONS OF VSD SYSTEM

3.1. Process Description

The principal difference between VSD and conventional ATAD systems is in its using an in-ground hyperbaric reactor. Installed by conventional drilling techniques, the VSD reactor is typically 110 m (350 ft) deep, occupying only a fraction of the area used by conventional surface digestion systems. The diameter of the reactor, which can range from 0.75 to 3 m (2.5–10 ft), is determined by the quantity of biosolids requiring treatment. While traditional above-ground ATAD processes use two or three tanks in series to achieve sufficient temperatures and prevent short-circuiting, VSD combines the stages within a single reactor.

As shown in Fig. 2, the VSD VERTAD reactor has three separate treatment zones: the oxidation zone, the mixing zone, and the lower plug-flow or soak zone. The oxidation zone is the upper portion of the reactor, and includes a central concentric draft tube for circulation. The mixing zone is immediately below the oxidation zone. Air required for bio-oxidation within the upper zone is injected into the mixing zone. The injected air also provides airlift circulation. The lower plug-flow zone is designed to prevent short circuiting and provides the high-temperature residence time required to kill pathogens such as *Salmonella* and fecal coliform, ensuring that the product meets class A biosolids requirements set forth by the US Environmental Protection Agency (US EPA) in CFR-503 (Appendix 1).

3.2. Process Operation

Figure 2 shows the flow diagram of a complete VSD system including supplemental processes for pretreatment, thickening, dewatering and air emission control. The following shows how a VSD process is operated.

a. Screened sludge feed is delivered into the mixing zone where it is mixed with partially digested recirculating sludge.
b. Compressed air is continuously added below the mixing zone to provide the oxygen required by the microorganisms to digest the sludge. The high hydrostatic pressure ensures a high oxygen transfer rate (OTR).
c. Air bubbles rising up the outer annulus create circulation up the annulus, into the head tank, and down a central draft tube.
d. Off-gas containing excess air and carbon dioxide formed by microbial respiration disengages in the head tank and vents to an off-gas biofilter that effectively breaks any foam and removes odor.
e. A small fraction of the recirculating sludge moves from the mixing zone into the lower plug flow zone, which is designed to prevent short-circuiting. In this zone, residual organic materials are digested and the high temperature ensures that pathogens are destroyed.
f. Class A biosolids are withdrawn from the bottom of the reactor through a central discharge pipe and transferred rapidly to a product tank at the surface.
g. The rapid depressurization of the digested class A biosolids causes the solids to separate in the product tank by flotation, and yields class A biosolids prethickened to around 10% solids. The subnatant liquid is recycled back to the sewage treatment plant for processing before discharge.

3.3. *Process Applications*

As stated previously, VSD is an advanced autothermal thermophilic aerobic digestion (ATAD) process. This technology uses a subsurface vertical reactor to aerobically digest mixed primary and secondary biosolids. Enhanced oxygen transfer in the process facilitates high metabolic activity resulting in heat generation. This enables the production of class A biosolids at short solids retention times (SRT).

The VSD digestion system is commercially known as VERTAD™, manufactured by NORAM Engineering and Constructors Ltd., (Vancouver, BC). Unlike conventional ATAD processes, the state-of-the-art VSD aerobic thermophilic process converts municipal primary and secondary sludges to class A biosolids. It uses an in-ground hyperbaric aeration reactor—a device that has been proven effective through more than 20 yr of commercial operation in biological treatment processes. The VSD reactor's patented design, according to its manufacturer, has the following advantages over conventional ATAD:

a. Excellent volatile solids destruction (>40% in a 4 d hydraulic residence time, or HRT).
b. Produces class "A" biosolids product (40 CFR 503.32, Alternative 1).
c. Flotation thickening using dissolved gases in the product.
d. Thickened product dewaters to high solids content with low polymer demand.
e. Efficient space utilization because of to its minimal plant footprint.
f. Highly efficient oxygen transfer.
g. Low volumes of process air to treat in subsequent off-gas biofilters.
h. Power costs are substantially lower than conventional aeration processes.
i. Enhanced microbial degradation because of efficient, high energy mixing.
j. Autothermal operation produces heat that is available for recovery.
k. Constructed using conventional well drilling or mining techniques.
l. Simple open-pipe aeration device requires very little maintenance.
m. Odor, volatile organic compounds (VOC), and ammonia emissions are minimal compared with conventional processes.
n. Off-gas from head tank is contained and easily routed for biofilter treatment.
o. Lower capital cost than conventional class A technologies.

p. The system can be economically enclosed in a building in locations where climatic conditions are unfavorable or if it is desirable for the plant to architecturally blend in with the surrounding environment.
q. The system is uncomplicated, easy-to-operate and maintain, and well-suited to fully-automated unattended operation.
r. The in-ground reactor is much less likely to sustain damage in an earthquake than above-ground reactors.

The VSD system is ideal for treating biosolids streams from a VSB system, or from a conventional biological treatment plant treating municipal or industrial wastewater. It has particular advantages in applications with the following conditions:

a. Sites with high biosolids disposal and/or trucking costs.
b. Applications in which class A biosolids are required.
c. Sites with space constraints.
d. Retrofits and plant expansions.
e. Sites with high precipitation or extreme temperatures.
f. Sites close to residential areas.
g. Locations where large unsightly plants are undesirable (i.e., recreation areas).
h. Sites in areas with high seismic activity.

4. DESIGN CONSIDERATIONS OF A COMPLETE VSD SYSTEM

A complete VSD system includes not only the main process digestion unit, but also the supplemental units, such as flotation thickener, supplemental anaerobic digester, biosolids dewatering unit, and air emission control system. The combination of the above units together will accomplish the objectives of biosolids stabilization, biosolids dewatering, supernatant recycling, and air emission control.

4.1. ATAD-Air

ATAD-air is a form of advanced aerobic digestion that operates in the thermophilic temperature range (>45°C) using air as the source of the required oxygen. The operation is autothermal, i.e., 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. The digesters are covered and insulated to minimize heat losses from the system. Use of oxygen in place of air (ATAD-oxygen) is another similar advanced autothermal thermophilic aerobic digestion process, which is introduced in the next section. VSD can be either a VSD–ATAD-air, or a VSD–ATAD-oxygen, involving the use of vertical shaft reactor.

This section introduces only ATAD-air and VSD–ATAD-air. Both processes share the same theory and principles, except that VSD adopts a vertical shaft reactor instead of an above-ground tank. All design criteria developed for conventional ATAD-air can be applied to VSD-air as well. Since 1977, one full-scale ATAD-air unit has been operated at the Binghamton-Johnson City, New York wastewater treatment plant. Engineering results have fully demonstrated the feasibility of this process and have provided the technical knowledge is presented next.

ATAD–air and VSD–ATAD-air can be applied to biosolids with solids concentrations of 1.5% or more. More dilute biosolids will not reach thermophilic temperatures without supplemental heat. The high temperatures reached in the digester might result in virtually complete destruction of pathogens and eliminate the need for further disinfection. 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 reduction can be achieved with less than 8 d 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 might improve the settle-ability and dewatering characteristics of sludge. The simplicity of operation might be suitable for use in small treatment plants. It could also have application in cold climates where conventional aerobic digestion is ineffective or requires excessively long detention times.

The ATAD-air process is not applicable to conventional waste-activated sludges (WAS) because of the large amount of heat required to raise WAS (at 0.5% solids) to thermophilic temperatures. The process has high operating costs, primarily for air supply. The oxygen transfer efficiencies required to maintain thermophilic conditions with air might be as high as 15%. To achieve the high oxygen transfer efficiencies required, the system used was proprietary in nature; the "Liacom System" by DeLaval Inc., which utilized a self-aspirating aerator. The VSD system (VARTAD™) marketed by NORAM, uses the deep shaft reactor as well as covers and jacketing to contain the heat.

Based on full scale ATAD-Air system studies, some selected parameters for a conventional (nonvertical shaft) 1000 ft^3 reactor are as follows (42):

a. Retention time (5.4–7.7 d).
b. TVS loading rate (0.17–0.26 $lb/ft^3/d$).
c. Treatment efficiency (TVS removal 22.1–37.2%).
d. pH feed sludge (5.4–6.1).
e. pH reactor (7.6–7.9).
f. pH effluent (7.6).
g. Ambient temperature (15–25°C).
h. Biosolids feed temperature (20°C).
i. Reactor temperature (48–52°C).
j. Oxygen transfer efficiency (8.7–15.1%).
k. Air flow (0.78–0.91 ft^3/s).

Generally, air adjustment, pH adjustment and mechanical foam cutting are required. Residuals generated include both the supernatant and the digested biosolids. General design criteria are: reactor temperature 45–70°C and retention time 2–10 d. The full-scale US EPA demonstration project indicated very few problems with the ATAD-air process or equipment reliability. During winter conditions (ambient: –20°C) the digester remained in the thermophilic range. There were no operational problems with the self-aspirating aerator system. Generally, there are indications that the ATAD-air process is more stable than anaerobic digestion and more easily able to recover from extreme conditions.

When vertical shaft reactors are used instead of conventional above-ground reactors, both oxygen transfer efficiency and treatment efficiency are higher and detention time is shorter. VSD–ATAD-air requires less space than conventional digestion and, by stabilizing and disinfecting the biosolids, reduces the adverse impact of land application of biosolids.

4.2. ATAD-Oxygen

Autothermal thermophilic aerobic digestion using pure oxygen (ATAD-oxygen and VSD–ATAD-oxygen) is a form of another advanced aerobic digestion that operates in the thermophilic (>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 digesters are covered to minimize heat losses from the system. Heat loses are also reduced in pure oxygen systems because there is little exhaust gas to remove the heat generated by the process. The equipment for pure oxygen thermophilic aerobic digestion (ATAD-oxygen and VSD–ATAD-oxygen) is similar to that of the other advanced aerobic digestion (ATAD-air and VSD–ATAD-air) discussed previously with the addition of digester covers and an oxygen generator.

Since 1980, two full scale studies (Denver, Colorado and Speedway, Indiana) have been conducted using pure oxygen aerobic digestion. ATAD-oxygen and VSD–ATAD-oxygen systems might have the greatest applications where pure oxygen-activated sludge processes are used. The high temperatures used by the process might result in virtually complete destruction of pathogens, and eliminate the need for further disinfection. In colder climates the ATAD-oxygen and VSD–ATAD-oxygen processes 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 5 d retention. The high temperatures decrease oxygen requirements because of the inhibition of nitrification. In general, all aerobic digestion processes produce supernatants with lower organic loadings than anaerobic digestion. The danger of methane explosions is also reduced.

ATAD-oxygen process system might not be applicable to conventional unthickened waste-activated sludge (WAS) because of the large amount of heat required to raise WAS (at 0.5% solids) to thermophilic temperatures. The ATAD-oxygen 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 requirements and smaller reactor volume might enable the process to be competitive with other forms of digestion, particularly when a pathogen-free sludge is desired.

Table 1 presents the US EPA's performance data for ATAD-oxygen systems. The requirements of physical, chemical and biological aids, and the generation of residuals of ATAD-oxygen and VSD–ATAD-oxygen systems are the same as those of ATAD-air and VSD–ATAD-air systems.

When vertical shaft reactors are used instead of conventional tank reactors, oxygen transfer efficiency, and treatment efficiency will be higher, and detention time will be shorter. The design criteria for both single- and two-stage systems are similar: (a) retention time: 5 d or less, and (b) reactor temperature: 45–60°C. The ATAD-oxygen process, such as VSD, is stable and can more easily recover from extremes than anaerobic digestion.

Table 1
Performance of Autothermal Thermophilic Aerobic Digestion Using Oxygen (ATAD–Oxygen)

Single stage system	Phase I	Phase IA	Phase II	Phase III
Sludge description	O_2 step feed	O_2 step feed	O_2 activated sludge	Primary + O_2 activated sludge
Temperature (°C)	14–18	17–19	17.4–22	16–22
pH	6.0–6.3	5.9–6.4	5.9–6.4	5.5–6.1
TSS (mg/L)	25,000–33,000	30,000–34,000	25,000–40,000	–
VSS (mg/L)	21,000–27,000	22,000–27,000	20,000–30,000	–
TS (mg/L)	–	–	–	30,000–49,000
TVS (mg/L)	–	–	–	22,000–35,000
Retention time (d)	4.2	4.2	4.2	4
Digester temperature (°C)	47.3	46.4	50.4	50.2
VSS loading rate (lb/ft^3/d)	0.36	0.38	0.37	0.45
VSS reduction (%)	37	30	40	30

Two stage system (multiple test runs combined)	O_2–WAS	Primary + secondary sludge
Temperature (°C)	12–24	12–30
pH	5.9–6.9	6.0–6.6
TS (mg/L)	26,000–50,000	23,000–60,000
TVS (mg/L)	18,000–38,000	18,000–41,000
Retention time (d)	3.7–5	3–5
Digester temperature (°C)	48.7–57.8	45.3–52
VS loading rate (lb/ft^3/d)	0.32–0.46	0.38–0.53
Overall VSS reduction (%)	29–42	30–45

4.3. *Flotation Thickening After VSD*

In conventional DAF systems, recycled subnatant flow is pressurized to 30–70 lb/in^2 gage (psig) and then saturated with air in a pressure tank. The pressurized effluent is then mixed with the influent sludge and subsequently released into the flotation tank (12). However, the flotation thickener after VSD is not a conventional DAF, because the VSD effluent has already been pressurized in the reactor to a high pressure (350 ft of hydraulic water head). The VSD effluent containing supersaturated gas can be directly released in a flotation tank for biosolids thickening. This new flotation process is called VSF, a process that does not need the conventional DAF pressure tank (gas dissolving tank or gas dissolving tube). The excess dissolved gas in the VSD effluent separates from solution at the atmospheric pressure in the vertical shaft flotation (VSF) thickener. The minute, 80 μm, rising gas bubbles attach themselves to biosolids particles which form the floating sludge blanket at the water surface. The floating thickened biosolids are skimmed off and pumped to the downstream biosolids handling facilities whereas the subnatant is returned to the plant's headworks. Polyelectrolytes are frequently

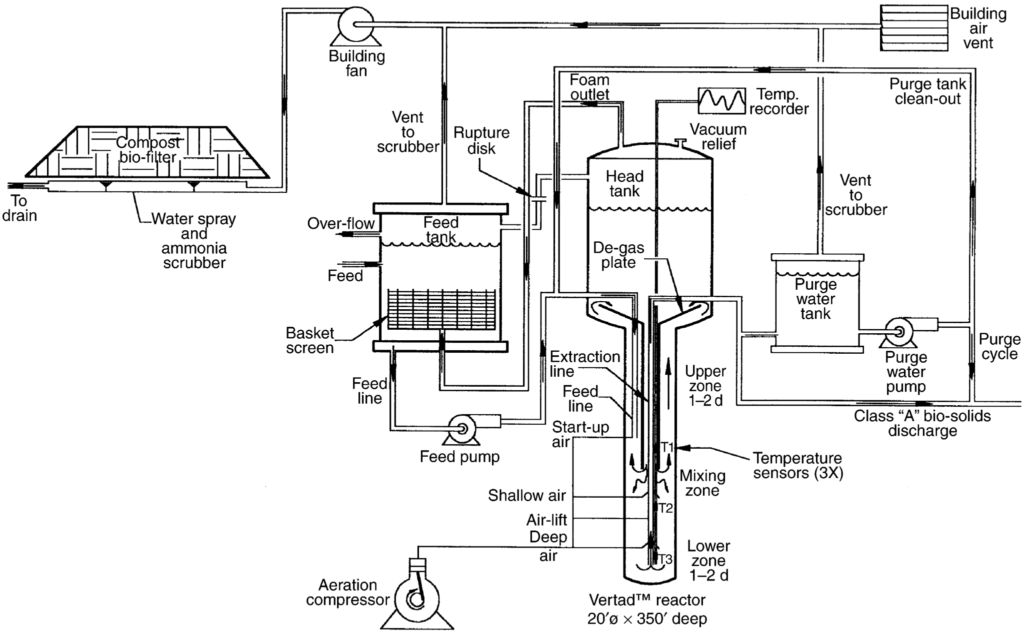


Fig. 3. VSD (VERTAD™) demonstration facility process flow diagram.

used as flotation aids to enhance performance and create a thicker biosolids blanket. A flow diagram of the VSD-VSF process system is shown in Fig. 3.

Although VSF is a new technology, yet it is similar to DAF in efficiency, theory and principles. DAF is the most common form in the United States for the thickening of WAS, and to a lesser degree combined sludges. DAF also has widespread industrial wastewater applications. It is expected that VSF will compete with DAF favorably from both technical and economical view points. The use of VSF is limited primarily to the thickening of biosolids following VSD and before dewatering or anaerobic digestion. Used in this way, the efficiency of the subsequent dewatering units can be increased and the volume of resulting supernatant is decreased. Existing VSF thickening units can be upgraded by the optimization of process variables and the utilization of polyelectrolytes.

With VSF thickening, it is possible to attain biosolids concentrations of up to 6% compared with a maximum of 2–3% that can be achieved for WAS in gravity thickening. Data from various flotation thickening units indicates that solids recovery ranges between 83 and 99% at solids loading rates of 7–48 lb/ft^2/d. Flotation aids, mostly polyelectrolytes, are commonly used to enhance performance.

VSF thickening requires less land area than gravity thickeners. The subnatant stream is returned to the head of the treatment plant. The gas released to the atmosphere may strip volatile organic material from the biosolids. The volume of sludge requiring ultimate disposal might be reduced, although its composition will be altered if chemical flotation aids are used.

4.4. Optional Dual Digestion System

The process of dual digestion involves the use of an aerobic digestion process as a pretreatment step before mesophilic anaerobic digestion. Dual digestion is a well established class A biosolids process. The majority of current systems utilize an aerobic digestion step with a short contact time (around 24 h) and pure oxygen to support the biological process (28). There are two possible approaches to dual digestion that involve operating the aerobic first stage at different retention times. In one scheme the aerobic stage is operated at an HRT of 1–2 d to achieve class A pathogen removal and some level of stabilization (10–20% volatile solids (VS) removal). A second approach involves a longer HRT in aerobic digestion (between 4–6 d) which would achieve class A pathogen removal and a more significant level of stabilization (35–50% VS removal).

In VSD systems (either VSD–ATAD-air or VSD–ATAD-oxygen), organic nitrogen, and fat, oil, and grease (FOG) are preferentially degraded over organic solids primarily consists of cellulose. This is significant when considering a dual digestion flow sheet with VSD pretreatment ahead of anaerobic digestion. The technologies are complementary in that the VSD systems can readily degrades fats and proteins, compounds known to cause scum build-up and mixing problems in mesophilic anaerobic digesters, and the anaerobic digestion process is capable of destroying the cellulose material typically present in VSD product (3).

Using a dual digestion system will result in increased overall volatile solids destruction compared with that of either VSD or anaerobic digestion alone. Systems using dual digestion can achieve as high as 70% VS destruction in a 15 d solids residence time (SRT) (3,27). This compares favorably to VSD systems which achieve a VS destruction of 40% in a 4 d SRT, and mesophilic anaerobic digestion which can achieve 50% VS destruction in a 20 d SRT. Although the VSD process does not generate methane gas, it does produce recoverable heat in the form of hot water, which can be used to heat the mesophilic digesters in a dual digestion system. The methane produced in anaerobic digestion can then be used for other purposes.

The majority of the publically owned treatment works (POTWs) in North America still utilize mesophilic digestion as the sole process for stabilizing sludge. This process successfully produces methane gas as well as a stabilized class B biosolids product. The impetus for facilities to explore combined digestion is the fact that many of these municipal treatment facilities generate significant quantities of biological solids, and need to maximize solids destruction in order to minimize solids handling and disposal costs (3). Retrofitting an existing mesophilic anaerobic digestion system to a dual digestion system offers the following benefits:

a. Production of class A biosolids.
b. Increased overall volatile solids destruction (as high as 70% VS destruction in a 15 d SRT compared with 50% VS destruction in a 20 d SRT in mesophilic digestion alone).
c. Heat recovered from the ATAD process can be utilized for sludge conditioning, as well as building and anaerobic digester heating.
d. Methane is produced in anaerobic digestion, and the overall bioenergy recovery from the dual digestion system is higher than that from aerobic or anaerobic alone.
e. Improved operation in the mesophilic anaerobic digestion stage (improved mixing, less scum, operation at higher solids concentrations).
f. Significantly reduces the size (or increases the capacity) of the dewatering system, improved dewaterability.
g. Reduces recycle nutrient loading to the treatment facility.

4.5. Biosolids Dewatering Processes

Dewatering is the removal of water from biosolids to achieve a volume reduction more than that achieved by thickening. Dewatering of biosolids is desirable for one or more of the, following reasons:

a. To prepare sludge for landfilling.
b. To reduce sludge volume and mass for lower transportation costs.
c. To reduce the moisture content and thereby increase the net heating value to make incineration more economical.

Some dewatering processes use natural means, such as evaporation, percolation, and so on (43) for moisture removal; others use mechanical devices to speed the process. The method chosen for dewatering is determined mainly by the type of biosolids, space available, subsequent processes, and economics.

The most common biosolids dewatering methods are as follows:

a. Vacuum filtration.
b. Filter press.
c. Belt filter.
d. Centrifugation.
e. Thermal drying.
f. Drying beds.
g. Lagoons.

All of these technologies are well established. The most commonly used methods for dewatering industrial biosolids are lagoons and drying beds. The mechanical methods for dewatering biosolids using vacuum filtration, centrifuges, and filter presses are also in widespread use.

All biosolids dewatering processes except drying beds and lagoons are complex mechanical systems. Their reliability is, thus, dependent on operator skill and proper maintenance. Vacuum filtration requires considerable operating attention and proper chemical conditioning to prevent filter blinding. Filter and belt filter presses have several moving parts and require maintenance to obtain a high level of reliability. Centrifuges are high speed mechanical devices subject to maintenance problems.

The performance of dewatering devices is measured by biosolids concentration or cake moisture, and solids recovery. Dewatered concentrations of 10–30% solids are common with biosolids and values of 60% solids or more may be attained with some inorganic residues. The performance of any one specific dewatering method depends on biosolids type, characteristics, conditioning, and operating conditions.

4.6. Gas-Phase Biofiltration for Air Emission Control

A VSD system is similar to other aerobic digestion processes; therefore, an air emission control unit for removal of odor and VOCs is not absolutely required. Occasionally, the digester influent does contain toxic VOCs and odorous substances, and the aerobic digester has to be enclosed for air emission control. Under this adverse condition where enclosure is required, conventional aerobic digesters and conventional autothermal thermophilic aerobic digesters will not be economically feasible because of their big foot prints. The construction costs of an enclosure for collection and subsequent treatment

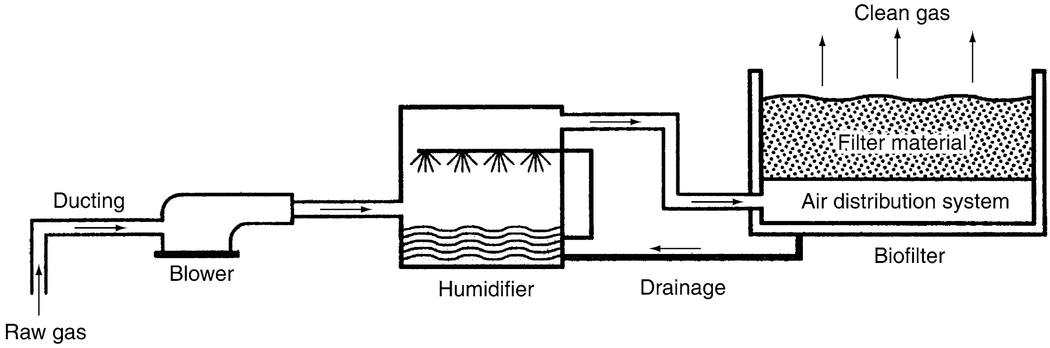


Fig. 4. Biofiltration flow diagram.

of polluted air emission streams will be extremely high. VSDs with typical diameters in the range of 2.5–10 ft, would have small foot prints that will allow VSDs to be economically enclosed for air emission. This section introduces one of many air emission control technologies which can be applied in conjunction with VSD units. Biofiltration frequently teams up with VSD because of its low cost. Air emission control technologies and their related costs can be found in "*Air Pollution Control Engineering*" and "*Advanced Air and Noise Pollution Control*" refs. 15,44.

4.6.1. Biofiltration Process Description

Biofiltration is an emerging technology for controlling VOCs emission in waste gas streams. Biofiltration has been extensively used in Europe, especially for odor control and it has been demonstrated at full-scale in the United States (14). In the biofiltration process, the waste gas is vented through a biologically active material where the biodegradable VOCs are oxidized into carbon dioxide and water. Physical sorption and chemical degradation might also occur and contribute to the overall removal efficiency. Figure 4 is a schematic diagram of a typical single-bed biofilter system. Because biofilters are biologically sensitive, the temperature and moisture of the gas and filter bed are extremely important in design considerations. Radial blowers are used to transport the waste gas to a humidifier. The humidifier saturates the gas stream to 95% relative humidity, which prevents drying out of the filter material. The effect of the filter drying out is death of the microorganisms and a resultant loss of control efficiency.

The gas stream enters the gas distribution system below the filter. As the gas diffuses through the filter, air contaminants will diffuse into the wet, biologically active layer (biofilm) where degradation occurs. Clean gas diffuses out the top of the filter. Excess drainage from the filter bed is the only potential source of wastewater discharge. In particular, where drainage contains regulated organic contaminants, the drainage is recycled to the humidifier to minimize wastewater discharge. Because particulates in the waste stream might clog the humidifier and the biofilter, a pie-filter might be required. A heat exchanger may also be required to heat or cool the waste gas stream if temperatures are not within the optimum range (20–40°C).

Typically, the filter material is compost, peat, wood chips, or soil with an inert material such as polystyrene particles. As the VOCs are degraded, water, carbon dioxide, mineral salts, and biomass are generated. Mineralization leads to compaction of the filter material, which causes an increase in backpressure. Typically, the filter material is turned over after 2 yr of operation and usually replaced 1–2 yr after turning over the filter to prevent backpressure problems (45). The most common biofilter system is an open, single-bed system. The clean gas is vented directly to the atmosphere in an open biofilter. Enclosed, multiple-bed systems can be stacked and have been used for low maintenance and space constraint situations.

4.6.2. Applicability to Air Emission Control

The applicability of biofiltration is dependent on the characteristics of the waste gas emitted from a VSD or other treatment process. Typical biodegradable contaminants include: alcohols, ethers, aldehydes, ketones, amines, sulfides, and certain monocyclic aromatics (xylene, benzene, toluene, and phenol). Waste streams containing chlorinated solvents are not readily biodegradable and are not appropriate for emissions control by biofiltration.

Biofiltration, as a VOC control technology, results in the complete degradation of the biodegradable contaminants and avoids the cross media transfer of pollutants. A major requirement, and thus limitation, of biofiltration is the absence of biologically toxic substances in the waste gas, such as heavy metals. The technology is limited to biodegradable components.

Because biofiltration is biologically sensitive, the potential system failures represent areas that should be considered when evaluating this technology. An undersized filter can result in VOC air emissions because of insufficient treatment. As the filter is sized by off-gas flow rate and concentration, the off-gas should remain within these design parameters during operation to prevent the loss of control efficiency. Inadequate preconditioning of the off-gas for temperature, moisture, particulates, or toxic constituents can also result in the complete loss of control efficiency.

Intermittent off-gas streams can be treated with a biofilter assuming the flow rate and concentration of the gas stream are within the design values. Filter beds can survive shut down periods of at least 2 wk without any significant reduction in biological activity. Shut down period up to 2 mo is feasible with nutrient addition and aeration of the filter (14,45). Biofiltration is technically and economically feasible for controlling VOCs in large volume gas streams with low concentrations. One potential use of biofiltration is odor control at POTW sites assuming the odor constituents are biodegradable. Usually, because odor problems are caused by compounds with low odor thresholds, off-gas concentrations often will be relatively low.

4.6.3. Range of Effectiveness

Usually, biofiltration is cost effective for large volume gas streams with relatively low concentrations (<1000 ppm as methane) of easily biodegradable contaminants (14). Maximum influent VOC concentrations have been found to be 3000–5000 mg/m^3 (45). For optimum efficiency, the waste gas should be 20–40°C and 95% relative humidity. The filter material should remain at 40–60% moisture by weight and have a pH between 7.0 and 8.0. For most easily biodegradable constituents, control efficiencies more than

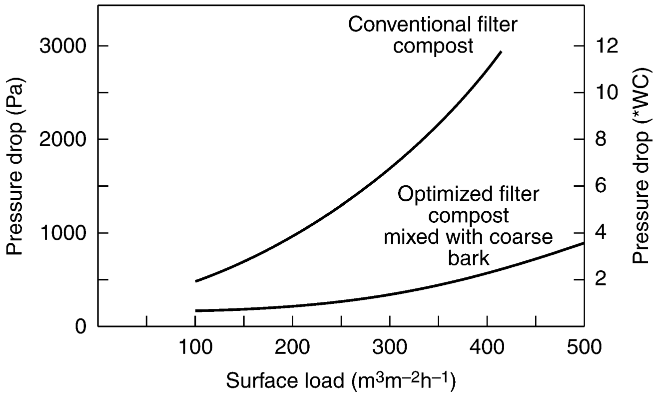


Fig. 5. Pressure drop of two biofiltration systems as a function of surface loading rate.

90% are achievable (14). Degradation rates for common air pollutants are typically from 10 to 100 g/m^{3}h (45). The key parameters affecting the control efficiency of a biofiltration system include the environmental conditions in the filter material, biofilter design, filter size, and waste gas composition. The filter must also have a large reactive area and low pressure drops; therefore, compaction must be kept to a minimum.

4.6.4. Sizing Criteria of Biofiltration

Typical biofilter systems have been designed to treat 1000–150,000 m^3/h waste gas with the systems having 10–2000 m^2 of filter area (14). The depth of biofilter material is typically three to four feet. The size of a biofilter system is dependent on the following parameters:

a. The loading rate of waste gas.
b. The concentration of compounds in the waste gas.
c. The rate of degradation of the compounds per unit volume.

Surface loads up to 300 m^3/h of waste gas/m^2 filter area is feasible without excessively high back pressures (45). The type of filter material affects the pressure drop across the filter. The effect of filter material on pressure drop is shown in Fig. 5 as a function of the surface loading rate.

The filter's large mass often provides sufficient buffer capacity to prevent breakthroughs during peak loadings, which allows sizing based on average hourly peak loads (14). The removal process in biofilters has been postulated to be controlled initially by a first-order-type biodegradation rate, but to be limited by transport properties at low inlet air flow rates (14). Usually, pilot testing of industrial waste gas streams with multiple contaminants is required, rather than modeling, to accurately size the full scale system.

4.6.5. Cost Estimating Procedure

Capital costs have been estimated at 77–123 USD/ft^2 filter area for installed open, single-bed biofilter systems. Costs of open, multiple-bed systems are approximately two times these costs. Enclosed systems have been estimated to cost between 123 and 677 USD/ft^2 filter area, depending on the size of the biofilter and the degree of process control (45). Operating costs are 0.45–2.06 USD/100,000 standard ft^3, not including filter replacement costs (14). Maintenance costs are about 1 h/m^2 of filter/yr.

Cost estimates were updated from 1991 to reflect the 2004 costs using the Cost Index for Utilities (Appendix 2); all costs were multiplied by a factor of 506.13/392.35 = 1.29 (46).

4.7. Operational Controls of Biofiltration

Operational controls are those procedures or practices inherent to the operation (and design) of control systems that can be followed to minimize the overall long-term emissions. Among these are:

a. Adequate system design and installation.
b. Startup testing.
c. Preparation of standard operating procedures for operators.
d. Control of operating variables to minimize emissions.
e. Monitoring of system performance.
f. Minimization of process upsets and startups.
g. Preventative and routine maintenance.

Obviously, a properly designed and operated control system is necessary to achieve the required air emission control efficiency or air emission limits. The use of experienced contractors and vendors will help ensure that the system design and installation are done correctly. Startup testing is advisable, with as many test conditions examined, as possible and all meaningful data should be recorded and evaluated. Systematic checks of wiring, direction of fan and pump rotation, integrity (leak tightness), and so on, should be made. The startup testing results should be incorporated into the formal standard operating procedures prepared for and followed by the operators of the biofilter.

Operating variables can be controlled to minimize air emissions. The most obvious variable to control is the treatment rate; for example, the lower the feed rate to additional air emission equipment, the lower the mass of potential emissions. Other variables such as the aeration rate for biodegradation systems, also directly influence emissions. Controlling operating variables to minimize air emissions is not always straightforward. There might be a number of competing variables that must be balanced for optimal control system performance.

To properly operate control devices, the biofiltration system design and performance must be understood. Performance data can be generated by routine monitoring of influent and effluent emission levels, pressure drops, operating temperatures, and so on. Operators should maintain the monitoring system so that plugged lines, water in the lines, and so on, don't result in misleading readings. Proper maintenance is another obvious requirement for successful control system operation, including routine inspection of the equipment, and implementation of corrective action when needed.

5. CASE STUDY

A demonstration project was supported by the Technology Assessment Program of the King County (WA) Wastewater Treatment Division (47). This program was developed in 1991 to evaluate and test technologies to reduce the environmental impacts of treatment plant operations including the space required for solids handling, biosolids truck traffic, and odor emissions. The VSD technology was selected for evaluation based on the potential for a very small footprint, low odor emissions, and production of class A biosolids. A demonstration facility was constructed in 1998 and operated through 1999 at the sewage treatment plant (STP) in Renton, WA. Successful results in these tests have prompted the County to consider VSD as a retrofit for existing facilities, and for future projects (3).

The VSD demonstration project consisted of design, construction, and operation of a demonstration-scale, deep vertical reactor for thermophilic aerobic digestion located at the STP (Renton, WA). The project team led by E&A Environmental Consultants Inc. was responsible for the planning, design, construction and testing of the facility. The technology owner, NORAM Engineering and Constructors Ltd., was actively involved in all aspects of the test program. King County provided engineering, operations and maintenance support throughout the project. The facility was completed in January, 1998 and testing was completed in December, 1999.

The test program was based on the following objectives:

a. To evaluate the SRT and temperature requirement for compliance with the Vector Attraction Reduction and class A pathogen requirements of the US EPA 40 CFR 503.32 Alternative 1.
b. Evaluate reactor hydraulics, oxygen transfer efficiency (OTE), and energy balance.
c. Determine the dewaterability of the VSD effluent (cake solids, polymer demand).
d. Evaluate "dual-digestion" VSD as pretreatment for mesophilic anaerobic digestion.
e. Perform an economic analysis of the technology.

5.1. Facility Design and Construction

The demonstration facility is located at the STP operated by King County in Renton, WA. The wastewater treatment plant is a 115 MGD facility with primary clarification, activated sludge secondary treatment, co-thickening of primary and secondary solids by DAF, anaerobic digestion, and belt press dewatering (3,47). A summary of the design parameters for the demonstration facility is provided in Table 2.

The main component of the VSD facility is a 50 cm diameter, 107 m deep (20 in. × 350 ft) subsurface, vertical reactor. The reactor tube was placed by conventional drilling technology using the dual air rotary drilling method. Subsurface geology consisted of 50 m of coarse sand and gravel alluvium above bedrock of siltstone and shale. There were indications of flowing groundwater above the bedrock. Before project initiation, the county conducted an assessment of the potential for earthquake damage to a deep reactor. The study concluded that damage to the reactor likely would be less than that to surface tankage (48). This finding is consistent with similar studies that have been carried out for Pacific Rim installations including Japan, Alaska, British Columbia, and California.

The vertical reactor has three separate treatment zones. A diagram of the process illustrating these three treatment zones is shown in Fig. 3. The upper zone of the shaft (surface to 44 m depth) contains a central concentric draft tube for circulation. The shallow aeration header introduces compressed air below the draft tube to induce flow up the annular space and down the draft tube. Thickened solids (THS) are introduced into this completely mixed zone. The lower zone extends below the draft tube down to the deep aeration header (44–96 m depth). High OTRs are attained in this zone under pressures of 5–10 atm. Mixing between the upper and lower zones occurs gradually over several hours. An unaerated plug-flow zone extends below the deep aeration header

Table 2
Design Parameters for VSD Demonstration Facility

Influent characteristics	Value
Influent	Thickened municipal biosolids (THS)
Loading	500–1500 lb solids/d (2500–7500 pop. equivalent)
Solids concentration (%)	6.5
VSS (%)	78–80
Primary sludge(%)	60
WAS (%)	40
Temperature (°C)	20–21
Biofilter treatment—building exhaust (cfm)	800
Biofilter loading rate (cfm/ft^2)	5
Feed rate	
@ 3 d HRT (gpd)	1770
@ 6 d HRT (gpd)	889
Equipment inventory	
VERTAD™ bioreactor	
Casing	1 @ 20 in. diameter/350 ft deep
Draft tube	1 @ 10 in. diameter/143 ft deep
Extraction	1 @ 3 in. diameter/347.5 ft deep
Reactor volume	
Total	740 ft^3
Liquid	710 ft^3
Vessels	
Feed tank	1 @ 60 in. diameter/72 in. high
Digestor head tank	1 @ 60 in. diameter/72 in. high
Purge water tank	1 @ 38 in. diameter/48 in. high
Mechanical	
Aeration compressor	87 scfm, 150 psi, 25 hp
Feed pump	1–10 gpm, 50 psi, 3 hp
Purge water pump	20 gpm, 26 ft TDH, 5 hp

to the bottom of the reactor (96–107 m depth). This zone is hydraulically separate from the aerated upper zones (as confirmed by tracer tests). Stabilized product is withdrawn using airlift through a 7.6 cm pipe that extends to within 0.5 m of the bottom of the reactor. The effluent is batch discharged at intervals sufficient to ensure strict adherence to the time/temperature requirements for pathogen destruction of class A biosolids.

The support equipment for the reactor includes a THS supply loop, a feed storage tank, a feed pump with variable frequency control, a purge water system, a 25 hp air compressor, a heat exchange system, a programmable logical controller (PLC), and a biofilter for off-gas treatment. The batch effluent withdrawal and feeding cycles (continuous or batch) are fully controlled by through PLC. Levels are continuously monitored by differential pressure sensors in the feed tank and reactor head tank. Temperatures are continuously monitored by sensors hanging at five elevations in a wet well in the center of the reactor. The THS supply loop provides a continuous supply of fresh undigested solids from the STP solids system storage tank.

The feed tank provides 2.2 m^3 of feed storage. Process air to the reactor is provided by a continuous duty, rotary screw compressor that requires 18.6 kW to produce 2.5 m^3/min at 1035 kPa (87 scfm at 150 psi). Compressed air is injected at 48 m and 96 m depths. The process air supplies oxygen for biological metabolism and induces mixing in the reactor. Air that is not dissolved produces voidage (volume of bubbles per unit volume of liquid) and is released from the reactor liquor in the head tank. A weighted check valve on the off-gas pipe provides up to 35 kPa (5 psi) back pressure to the head tank which reduces voidage. The off-gas is directed to the bottom of the feed tank to provide additional back pressure and capture foam and latent heat in the influent solids. The test facility is housed in a temporary building that is provided with utilities and air collection. Building exhaust and reactor process off-gas passes through a water scrubber for ammonia removal and is then processed through a biofilter.

A system to add supplemental heat to the reactor was installed after it became evident that heat loss from the pilot reactor exceeded the heat generated biologically and thermophilic temperatures could not be maintained. The reactor was not insulated and has a high surface area to volume ratio, which facilitates heat transfer to the environment. Also, flowing water was identified in three zones during drilling. Water moving past the reactor can remove substantial heat. To compensate for heat loss to the environment, reactor feed was initially preheated via steam injection using an 80,000 Btu/h propane-fired steam boiler. This was replaced later in the test period with a boiler (500,000 Btu/h) that supplied hot water to heat exchanger loops hanging in the reactor. This system provided direct control of the temperature in the reactor. The supplemental heating system was added rather than using a sludge–sludge heat exchanger to capture heat from the effluent. Whereas the VSD plant referenced is often cited as a pilot or demonstration scale facility, it should be noted that this plant can process the solids from a 7500 population equivalent. A 7500 population equivalent would be serviced by a 0.75 MGD VSD facility that would feed solids to a VSD plant of roughly this size. So while this plant is considered small by King County standards, it would be a full-scale facility for smaller municipalities.

5.2. VSD Demonstration Plan

The VSD demonstration program was designed to meet the goals of the King County research program to evaluate the viability of the technology with respect to reactor hydraulics, energy requirements, product quality, and the ability to meet the vector attraction reduction and pathogen destruction requirements of class A biosolids. An additional goal was to develop the design criteria necessary for full-scale design and cost evaluation. A range of operating conditions was tested in the facility. Before biological startup, cold water testing was conducted to evaluate reactor hydraulics and to test equipment. Next, preheating of the reactor using a hot water boiler provided data on heat loss to the environment in the absence of biological heat generation. Biological testing with varied HRT, temperature, aeration rates, and feed solids spanned the periods of January 15 to May 7, 1998, November 10 to December 17,

Table 3
Summary of Operating Ranges

Operating variable	Operating range
Hydraulic residence time (d)	2–6
Temperature (°C)	55–70
Aeration (scfm)	20–80
Feed solids content (%)	3.5–7

1998, and August 4 to December 23, 1999. Suspensions in operation between the various testing periods allowed ongoing modifications of the facility for improved data acquisition and control. During the third testing period, stable operation was achieved over a range of detention times and temperatures. In order to test a full range of conditions and determine the capabilities of the VSD process, some operating conditions were applied that did not provide a class A biosolids product. However, these imperative tests provided insights into the critical effects of such variables as sludge viscosity, oxygen transfer efficiency, and heat loss. The range of operating conditions that were tested for the process is summarized in Table 3.

The controlled parameters of the test program were the SRT, temperature, and aeration rate. The approach was to establish stable operations at specified operating conditions for a minimum of three detention times. During the fourth detention time, data was averaged to yield the reported values for the reactor performance under those stable conditions. Samples were collected for laboratory analysis of the THS feed, feed and head tank (upper zone) solids, and final effluent solids (from the deep extraction line). These samples were tested for total solids (TS), VS, pH, total Kjeldahl nitrogen (TKN), ammonia, chemical oxygen demand (COD), and alkalinity (ALK) by the STP laboratory according to Standard Methods. Fecal coliform and *Salmonella* analyses were conducted by the King County Environmental Laboratory. Additional laboratory and field testing included measurement of fat–oil–grease (FOG), total carbon (TC), total organic carbon (TOC), off-gas analysis, density testing, oxidation reduction potential (ORP), dewaterability, and dissolved oxygen (DO). Daily grab sample analyses of TS and VS were conducted whereas the remaining parameters were measured weekly. More frequent sampling and composite analyses were conducted during the fourth detention time. Temperatures, levels, and flows were logged and trended continuously through PLC using a Siemens WinCC trending program. Oxygen concentration in the off-gas was measured using a portable oxygen analyzer (first with a Quintox gas analyzer, and later a Teledyne Portable Flue Gas Oxygen Analyzer). The dewatering characteristics (polymer demand, cake solids content, and filtrate quality) of the digested product were tested by several dewatering equipment vendors (CIBA, US Filter, Andritz). Five gallon samples were delivered to vendor laboratories where testing was performed on bench scale centrifuges, belt presses, and capillary suction time (CST) test equipment. Onsite testing was conducted to compare VSD product to the mesophilically digested STP biosolids using jar testing to determine polymer demand, and press tests to assess the maximum achievable dry cake solids content.

Table 4
Volatile Solids Reduction at Varied Temperature and Residence Times

Test	HRT (d)	Temperature (°C)	Aeration rate (scfm)	VS reduction (%)
December 1998	4	56	56	40.9
September 1999	4	65	80	42.2
November 1999	3.4	56	36	42.3
December 1999	5.5	61	30	43.5

Tracer tests using both lithium chloride (LiCl) and table salt (NaCl) were conducted to assess the reactor hydraulic characteristics and to confirm that no short-circuiting was occurring in the reactor. LiCl or NaCl was batch loaded into the reactor and samples were collected from the reactor head tank and the product during batch product withdrawals. In the case of tracer tests involving LiCl, the samples collected were analyzed for Lithium content, and profiles were developed. For the salt traces (only performed in water), conductivity changes were measured in the samples taken from several depths in the reactor. Samples from lines at 213 and 268 ft depths were drawn continuously by a peristaltic pump at a rate of 1.3 L/min through 3/8 in. ID tubing weighted to keep it in place. Enough salt was added to increase the conductivity to approx 10 times the background concentration, ensuring good resolution. Conductivity was measured using the STP conductivity analyzer after proper temperature equilibration. This has automatic temperature compensation so readings need no further correction. Bench-scale testing was conducted at the University of Washington to assess the effect of VSD pretreatment on subsequent mesophilic anaerobic digestion (dual digestion). VSD product (4 d SRT) was fed to 3-L anaerobic digesters maintained at 11 and 15 d detention times. A control digester was fed STP-THS at an 11 d SRT. The digesters were maintained at 35°C. The main parameters used to evaluate digester performance included volatile solids destruction efficiency, gas production and percent methane, and product dewaterability using CST testing.

Odor panel testing was performed on samples collected from the VSD process. The odor panel analyses were conducted by Odor Science & Engineering, Inc., (OS&E). These tests were aimed at measuring the odor generated by the VSD process and the effectiveness of the biofilter for odor treatment. Odor was quantified by dilution-to-threshold (D/T) ratio and panelists described the odor character.

5.3. Design Criteria Development for VSD

5.3.1. Volatile Solids Destruction

A summary of the digestion performance results is presented in Table 4. The values reported are averages over a detention time after the process was stable for three detention times. A complete mass balance was achieved for each of these tests from which the reported efficiency values were calculated (3).

The effect of solids residence time on VS reduction was demonstrated by the testing. More than 40% VS reduction was demonstrated at a 4 d SRT. This efficiency appears to decrease approximately linearly as the residence time is reduced. In testing at a 2 d SRT and 67°C, a 21% VS reduction was demonstrated. As shown in Table 4, a 5.5 d SRT resulted in a 43.5% VS reduction. This value is considered conservative because concurrent testing of reactor response to oil and sugar addition complicated results because of the additional load on the system. Results from the three detention time conditioning period for the 5.5 d SRT test were averaging a 50.7% VS reduction before the supplemental additions. From these results it is believed that VS reduction will approach 50% with a detention time of approx 6 d at 60°C (360°C-d).

In general, it was found that an increase in temperature for a given solids retention time resulted in more VS reduction. Testing indicated that although temperature certainly affects biological activity, it is believed that the effects on water loss and oxygen transfer efficiency on reactor performance are more significant and important. Important findings about the effect of reactor sludge viscosity on oxygen transfer resulted in testing centered on controlling the reactor solids. With solids controlled at less than 4.5% TS, oxygen transfer efficiency nearly doubled, allowing a subsequent decrease in aeration rates. Decreased aeration rates minimized the amount of water loss (as latent heat) from the reactor for a given temperature. The difference between the December 1998 and November 1999 results can be explained by this finding. The two test periods were operated at a temperature of 56°C; however, the November 1999 trial was operated at a reduced SRT (3.4 d compared with 4 d), and at a reduced aeration rate (36 scfm compared with 56 scfm). The major difference between the two trials was that in the case of the November 1999 trial, the reactor solids concentration was being controlled at 3.5% TS, and in the December 1998 trial, the reactor solids concentration was 4.7%. Ultimately, the increased ability to transfer oxygen in to the mixture allowed a decreased SRT whereas simultaneously provided increased VS reduction.

The requirements for class A biosolids were met at an average detention time of 4 d at 60°C. As shown in Table 3, the system readily achieved more than 40% volatile solids destruction at varied detention times and temperatures. In order to satisfy the volatile solids destruction criteria of 38% (24) in conventional ATAD systems, Kelly et al. (29) suggested a 400°C-d product was necessary. The VSD results indicate that a 240°C d product exceed the US EPA requirements, with more than 40% volatile solids destruction.

5.3.2. Pathogen Destruction

Pathogen destruction was excellent with a 7 log reduction in fecal coliform and both fecal coliform and *Salmonella* below detection limits in the class A biosolids product (3). Fecal coliform and *Salmonella* were measured in the feed solids and digested VSD effluent weekly during the first operating period and intermittently during the third operating period. Fecal coliform in the feed solids averaged 5.39 E + 07 MPN/g dry solids and *Salmonella* averaged 5.87 MPN/4 g dry solids. Densities in the VSD effluent were consistently below the detection limit (fecal coliform: 5 MPN/g, *Salmonella*: 1.6 MPN/4g).

5.3.3. Reactor Mixing

The selected alternative for attaining class A pathogen control in the VSD process is by maintaining temperatures for the required contact time. Time and temperature requirements from the biosolids regulations (40 CFR 503) are shown in Fig. 6.

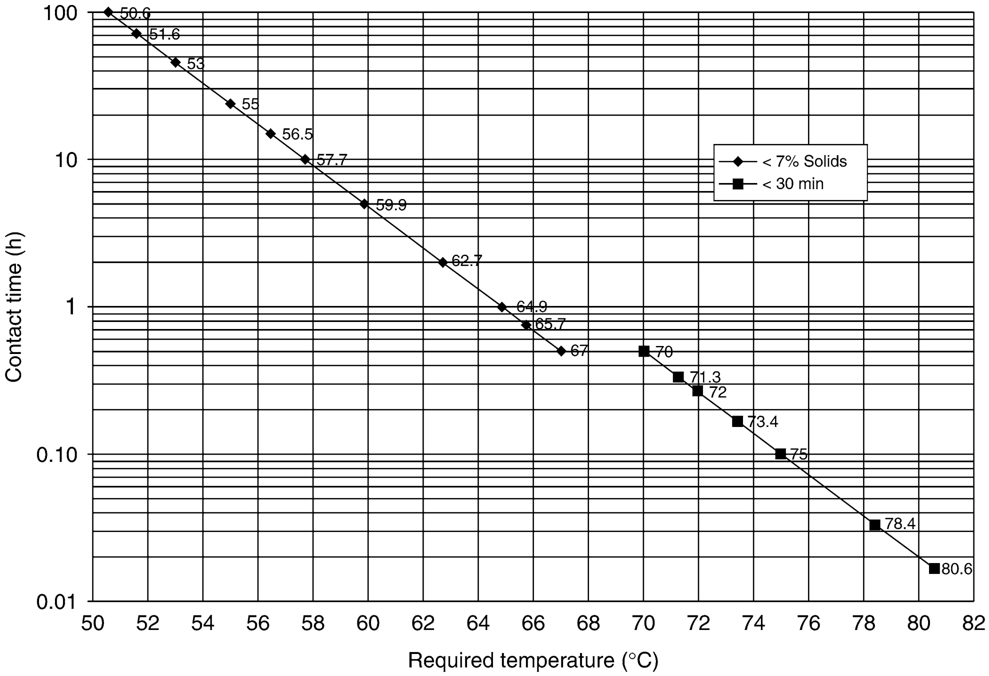


Fig. 6. US EPA 40 CFR 503 class A time and temperature requirements for solids less than 7%.

In order to test the reactor's compliance with time-temperature requirements, salt tracer studies were performed in the system. Samples were taken at regular intervals from four points in the system: the head tank (surface), 213 ft below grade surface (bgs), 268.5 ft bgs, and the deep extraction line. Critical distances in the system are: Upper aeration head - 158 ft bgs, lower aeration head - 315 ft bgs, and deep extraction line - 347 ft bgs. Head tank and intermediate sample points allowed observation of the saline dispersion as it moved through the reactor, providing an indication of the mixing time between the aeration headers. The deep extraction point allowed observation of the saline pulse, showing the time for a single particle to breakthrough the soak zone. A pulse of saline was pumped into the reactor quickly with enough salt for a 10-fold increase in reactor conductivity. After the pulse of saline, the system was fed and discharged continuously at a rate of approx 2 gpm (HRT of about 2 d). The conductivity profile vs time for the reactor tracer study is shown in Fig. 7 (3).

Tracer results are consistent with a model in which:

a. The upper zone (head tank to upper aeration head) is well mixed, with a time constant of the order of minutes.
b. The lower zone (upper aeration head to lower aeration head) is mixed gently by fluid rising in the wake of bubbles with a net turnover time which depends strongly upon air flow. In this study the lower aeration was 8 scfm, resulting in gentle mixing over approx 90 min. Here, simple theory based on the assumption that a bubble draws up its own volume of fluid are in reasonable agreement.
c. The soak zone is effectively plug flow.

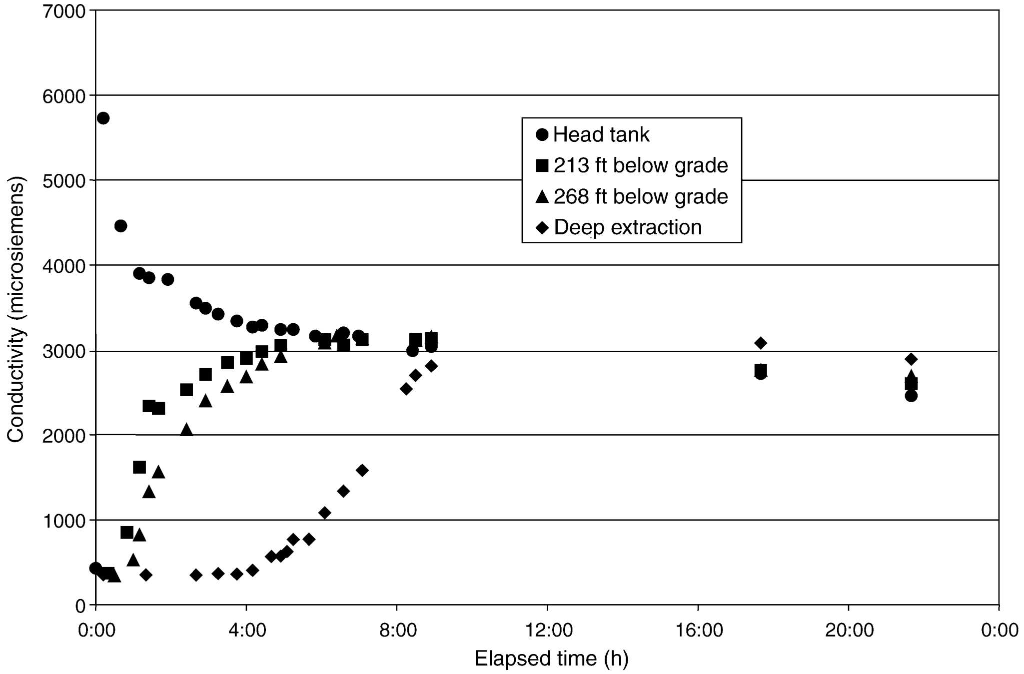


Fig. 7. VSD salt tracer study confirming no short circuiting.

The mixing test clearly indicates that the salt tracer did not reach the deep extraction point until approx 4 h had elapsed. This eliminates any concerns about short-circuiting in the reactor soak zone. The theoretical time for breakthrough (based on the 2 gpm extraction rate and the soak zone volume for plug flow) is 4 h 20 min. This is the first continuous feed, single reactor design that complies with the US EPA time-temperature regulations. Salt tracer studies confirmed that the VSD patented reactor design complies with time-temperature requirements (40 CFR 503 class A time and temperature requirements for solids less than 7%). These studies verify the true plug flow nature of the soak zone, and eliminate any concerns about short-circuiting in the system. Although it is believed that the demonstration facility's vertically stacked zone configuration complies with the time and temperature requirements, two variations are available to further assure compliance: (a) Installation of a flow restricting physical barrier between the slowly mixed and soak zones and (b) Maintaining a surface batch contact tank in which the VSD product is held for the required time at the appropriate temperature (using heat generated from the VSD).

5.3.4. VSF Thickening

During dewatering testing, it was indicated that the VSD effluent could be easily thickened after being discharged from the reactor. VSD effluent has the characteristic of high dissolved carbon dioxide concentrations because of the biological metabolism and the high pressure in the reactor. Acidifying the effluent (with sulfuric acid or alum) to

Table 5
Onsite Press Testing of VSD Product Dewaterability

Characteristics	Anaerobic	VERTAD™	VERTAD™ Acid float thickened
Cake solids (%)	20	32	31
Polymer dose (lb/t)*	17	17	17
Fitrate quality	Clear	Very turbid	Very clear

*1 t = 2000 lb.

approx a pH 5.0, releases the CO_2 as small bubbles, which attach to biosolids particles and float them to the compact surface blanket. Further testing resulted in float thickening of the VSD biosolids from 3.5% TS to 8–12% total solids, with a capture efficiency of approx 95%. Results were similar with both sulfuric acid and alum. Ferric chloride was not used, but it is expected to provide a similar result (3).

Analysis of the float or the thickened solids (THS) and the subnatant showed that nutrients partitioned into the digested biosolids. Thickened biosolids contained a phosphorus concentration 20–40 times the concentration in the subnatant. In the tests where sulfuric acid was used, the formation of ammonium sulfate caused the ammonia to get slightly partitioned into the biosolids. This result means that the phosphorus load that is typically recycled to the secondary treatment plant is being retained in the biosolids for beneficial reuse. The downstream implications of this flotation thickening step are as follows:

a. Significantly reduction in the size of the dewatering system.
b. Charge neutralization aids in dewatering.
c. Reduced recycle nutrient loading on the treatment facility.
d. Increased nutrient value of the biosolids.

5.3.5. Biosolids Dewatering

Test methods for dewatering included onsite press tests as well as outside laboratory testing at Andritz, IBA, and other vendors using bench scale belt presses and centrifuges. Samples tested included mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, VSD float thickened biosolids, and product from the combined VSD to anaerobic bench-scale test work (3,47).

Onsite press testing was performed using a set polymer dose of 17 lb/t for the mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, and the VSD float thickened biosolids. Cake solids were measured and the filtrate quality was reported qualitatively. The results are presented in Table 5.

Testing demonstrated that more than 30% cake solids could be attained with both the biosolids directly from the VSD reactor and the VSD float thickened biosolids whereas the anaerobically digested solids dewatered to 20% cake solids. In the case of the VSD reactor biosolids, the filtrate quality was poor, with losses of solids making the filtrate look very turbid. This indicated that a higher polymer dose would be required with the straight VSD product to obtain an acceptable filtrate quality. The VSD float thickened product outperformed both the anaerobic and VSD products. Not only did the VSD float

Table 6
Andritz Lab Centrifuge Testing of VSD Product Dewaterability

Characteristics	Anaerobic	VERTAD™	Acid float thickened VERTAD™
Cake solids (%)	12–14	31–34	31–34
Polymer dose (lb/t)	20.4	38	13.8
Capture efficiency (%)	95	96	99.5

thickened product have a very clear filtrate (clearer than that from the anaerobic dewatering tests); it obtained the best result from a cake solids perspective. This testing illustrated that the float thickening process highly enhances the dewaterability of VSD biosolids. Outside laboratory testing at Andritz, CIBA, and other vendors, was performed on samples of mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, and the VSD float thickened biosolids. Polymer dosing was optimized using 95% solids capture efficiency for the filtrate quality. Cake solids and solids capture efficiency were measured and reported. The results are presented in Table 6.

The test results showed that more than 30% cake solids could be attained with both the VSD reactor product and the acid float thickened product. Similar to onsite press testing, higher polymer doses were required for the VSD product withdrawn directly from the reactor (approximately double the polymer required for the anaerobic sludge). The anaerobic biosolids dewatered very poorly with the lab centrifuge, only attaining a maximum cake solids concentration of 14% (3). Like onsite press testing, the VSD acid float thickened product showed remarkable dewatering characteristics. It dewatered to high cake solids concentration (31–34%) with a lower polymer dose than that required for anaerobic sludge (13.8 and 20.4 lb/t, respectively). The conclusion is that the float thickening enhances the dewaterability of the VSD product. This is likely because of a charge neutralization that seems to act like a coagulant, aiding in dewatering. Generally, it is accepted that thermophilically digested aerobic biosolids can be dewatered to higher cake solids than anaerobically digested biosolids; however this has historically come with the expense of greater polymer demand. Murthy et al. (26) performed an examination of an autothermal process to isolate the cause of high polymer demand and high recycle chemical oxygen demand (COD). They found that the presence of monovalent ions in solution such as sodium, potassium, and ammonium ions can interfere with charge-bridging mechanisms occurring in the floc. This is a problem in conventional ATAD systems because the release of ammonium ions is the result of the absence of nitrification in the thermophilic process (25). This free ammonia release appears to be less pronounced in the VSD process, possibly because of the pressure in the reactor which results in the combination of free ammonia with dissolved CO_2, forming ammonium bicarbonate.

Murthy et al. (26) also found that the amount of biopolymer (proteins and polysaccharides) in solution was heavily correlated to increased polymer demand. They concluded that the concentration of biopolymers in solution was minimized by limiting the SRT of thermophilic digestion, and by minimizing the concentration of monovalent ions (specifically ammonia) in solution. These factors favor the VSD process because a

Table 7
Volatile Solids, FOG, and Organic Nitrogen Reduction

Test	HRT (d)	Temperature (°C)	VS reduction (%)	Org-N reduction (%)	FOG reduction (%)
December 1998	4	56	40.9	57.9	91.7
September 1999	4	65	42.2	49.8	80.8
November 1999	3.4	56	42.3	44.1	–
December 1999	5.5	61	43.5	49.9	80.4

relatively short SRT of 240°C-d is enabled by the high oxygen transfer achieved in the system, and ammonium bicarbonate is formed in the reactor, minimizing free ammonia.

5.3.6. Organic Nitrogen and FOG Destruction

A summary of the digestion performance results for VS, FOG, and organic nitrogen is presented in Table 7. The values reported are averages over one detention time after the process was stable for three detention times. A complete mass balance was achieved for each of these tests from which the reported efficiency values were calculated.

The reduction of organic nitrogen and fats, oils, and grease were relatively high considering the short SRT that the VSD process was tested at. The results were similar to the reduction efficiencies attained in the STP anaerobic digesters at a 28 d SRT. The organic nitrogen reduction was calculated based on the difference between total Kjeldahl nitrogen and ammonia in the feed and product. Generally, organic nitrogen reduction exceeded the total VS reduction. Analysis of the December 98 samples showed that protein degradation (assuming 6.25 kg protein/kg org-N) and FOG reduction accounted for 64% and 9%, respectively, of the VS reduction. The remaining VS reduction was attributed to carbohydrate reduction which is primarily consisting of cellulose and lignin. The preferential degradation of Org-N and FOG was further confirmed by visual inspection of the product which is very fibrous. Generally, these results are significant because undigested Org-N and FOG are responsible for the objectionable character of biosolids. These results also have significance when considering a dual digestion flow sheet with VSD pretreatment ahead of anaerobic digestion. The technologies appear to be complementary in that the VSD technology readily degrades fats and proteins, compounds known to cause scum buildup and mixing problems in anaerobic digesters, and the anaerobic digestion process is capable of destroying the cellulose material still present in the VSD product.

5.3.7. Biofiltration for Odor and Off-Gas Control

In the VSD system, the self-contained nature of the head works allows easy control over off-gas emissions. Off-gas can be easily routed to biofilters to remove the trace ammonia and dimethyl sulfide (DMS) compounds common with aerobic digestion technologies. Because of the high oxygen transfer efficiency in the bioreactor, the VSD process needs only a fraction of the air volume used in a conventional ATAD. As a result, significantly less off-gas is produced in the VSD process, reducing the size of biofilter required for off-gas treatment. Gaseous emissions from the VSD system are considerably smaller than those produced in conventional aeration processes. As mentioned previously,

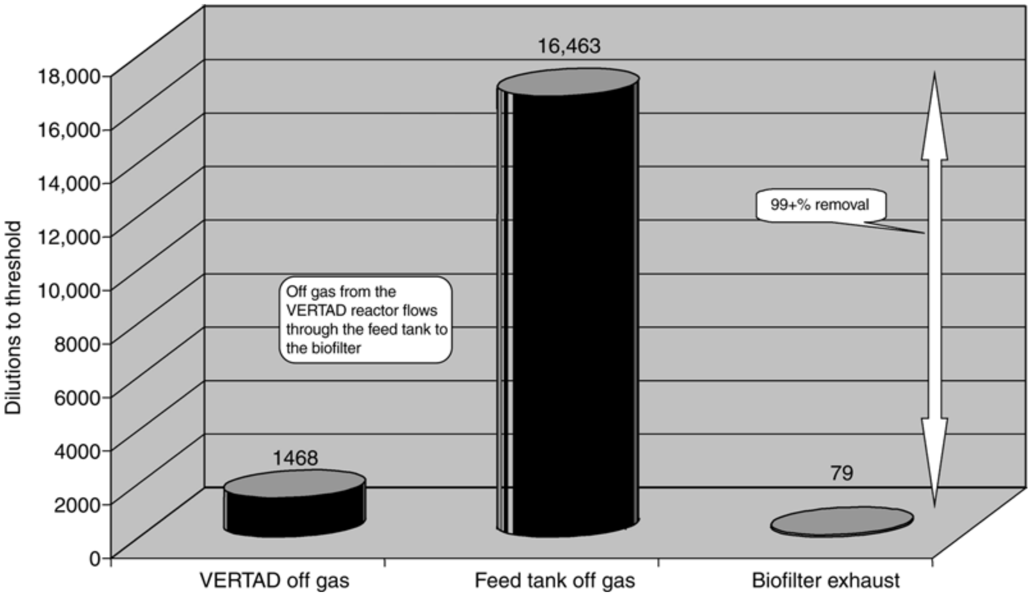


Fig. 8. Odor panel results of biofiltration process.

ammonia is converted to ammonium bicarbonate in the reactor, helping to eliminate ammonia emissions. In order to minimize the ammonia release from the system, reactors are operated at a maximum temperature of 60°C, preventing the dissociation of the ammonium bicarbonate.

Odor panel testing was performed on samples collected from the VSD process. The odor panel analyses were conducted by Odor Science & Engineering, Inc (OS&E). These tests measured the odors generated by the VSD process and the effectiveness of the biofilter for odor treatment. Odor was quantified by dilution-to-threshold (D/T) ratio and panelists described the odor character. The results of the odor panel work are provided in Fig. 8.

These results show that most of the VSD demonstration facility derived odor comes from the feed tank (16,463 D/T in 675 scfm) rather than the VSD reactor (1468 D/T in 36 scfm). The biofilter removed 99.5% of the odor loading (16,463 D/T in, and 79 D/T out). Odor panel testing has indicated that generally, the off-gas from the VSD process is odor-free. Character descriptors for the VSD off-gas before the feed tank included more terms such as compost, earthy, and vegetation. The off-gas from the untreated feed sludge tank changed the odor panel characterizations to focus on terms such as sludge and manure type odor. These results have highlighted the need to treat the off-gas directly from the reactor in a biofilter.

The reduced odor of the VSD off-gas is primarily attributed to the fact that the compounds primarily responsible for the objectionable character of unstabilized wastewater solids (FOG, Org-N) are the highest degraded fractions in VSD.

5.3.8. Oxygen Transfer Efficiency (OTE)

Oxygen transfer studies were performed to test the oxygen transfer rate (OTR) first into water and determine the theoretical maximum efficiency of the system, and second

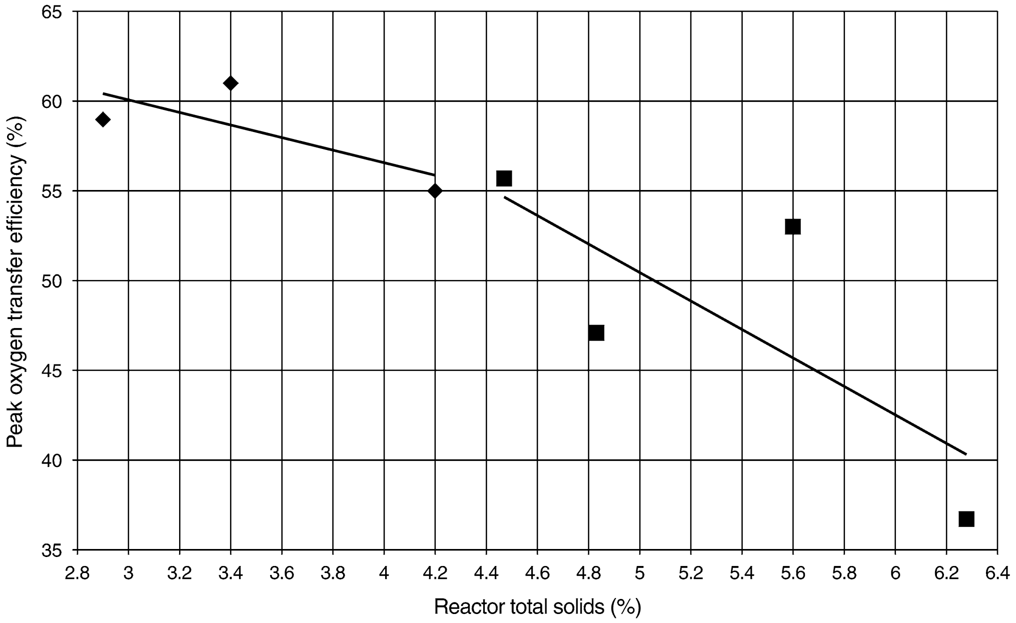


Fig. 9. Viscosity effects on peak oxygen transfer.

into sludge to determine the OTE attainable in the digestion process. The test method used to determine the OTR into water came from the American Society of Civil Engineers. This test involved the initial scavenging of dissolved oxygen with sodium sulfite and a cobalt chloride catalyst (Na_2SO_3 and $CoCl_2$), followed by reoxygenation to near the saturation level for the operating temperature. Throughout these tests DO was measured at multiple points, allowing the development of a mass transfer model. The OTR was measured for water, allowing calculation of the OTE in the system. The OTE was approx 66% into water at 54°C (129°F). This OTE represents a significant advancement in aeration technology over conventional aeration systems using air, which typically attain 10–20% OTE into water at 20°C, a lower temperature which facilitates oxygen transfer through increased solubility (3).

Sludge viscosity was found to have a pronounced effect on the OTE. As shown in Fig. 9, an OTE of 50% was attained easily at a reactor concentration less than 4.5% TS. At more than 4.5% TS, the transfer efficiency was diminished, as low as 35%.

Although sludge is highly non-Newtonian, and the concept of a Newtonian viscosity which is independent of shear rate is not strictly valid, some valuable order of magnitude generalizations can be made about transfer performance at higher VS destruction. In general, the OTE is improved at higher VS destruction because the viscosity of the bulk liquid is decreased with increased destruction, and decreased viscosity facilitates increased oxygen transfer. Transferring oxygen into thick sludge is not easy—even at high pressure, because of mass transfer limitations on the liquid side. Metcalf and Eddy (28) suggest that viscosity might decrease by a factor of two or more over the range of 3–6% for undigested sludge, with viscosity declining rapidly as sludge is digested. Doubling fluid viscosity changes oxygen diffusivity in the sludge in inverse proportion (i.e., it is reduced to half); the mass transfer coefficient and mass transfer rate will likely change by a similar order of magnitude. This is supported by the VSD OTE data which suggest that the OTE is nearly halved with a doubling in reactor solids, and that 4.5% is the practical operating cutoff before the sludge viscosity seriously affects the OTE. The effect of oxygen transfer upon heat release was corroborated during the oxygen transfer testing. During testing each test involving a lower aeration rate saw a systematic decrease in the reactor temperature. Each time the aeration rate was reduced, biological heat generation was reduced and a step change in temperature occurred.

The VSD system achieves very high oxygen transfer efficiency, more than 50% OTE can be expected when the viscosity of the reactor contents is controlled with a solids concentration less than 4.5% TS. These high OTRs are associated with the pressure and depth at which compressed air is introduced to the bioreactor. The high OTE results in enhanced digestion of the sludge and a decreased detention time to meet the class A biosolids requirements.

Generally, the OTE for other ATAD systems is not reported in literature presumably because of the proprietary nature of the systems; however, some independent data collected from an ATAD facility suggests that the VERTAD process compares favorably in terms of oxygen transfer efficiency. The VERTAD system achieves an average oxygen transfer efficiency of 50%, whereas a conventional ATAD system that was tested only achieved an average of approx 24% across a three stage system, presumably because of the high viscosity and low temperatures in early stages.

The increased oxygen transfer in the VSD system is thought to be the primary factor in decreasing the solids retention time to meet US EPA vector attraction requirements. As mentioned previously, Kelly et al. (29) have suggested that a 400°C-d product is necessary in ATADs to attain a volatile solids destruction of 38%. The VSD results indicate that a 240°C-d product exceed the US EPA requirements, with more than 40% volatile solids destruction. The difference in oxygen transfer and subsequent heat release in the two systems could explain this superiority of the VSD system.

5.3.9. Heat Balance

The small diameter of the demonstration reactor results in a large surface area to volume ratio, necessitating supplemental heat addition at the facility to maintain the required elevated temperature. A heat balance was performed using measured reactor heat loss data, influent and effluent temperatures, estimated biological heat production, aeration energy, and the measured supplemental heat necessary to maintain a set temperature. The heat balance showed that auto-thermophilic conditions would be maintained if the reactor diameter was increased to 0.8 m. (2.6 ft), thus, decreasing the relative surface area. Reactors of larger diameter will require a heat removal system to prevent overheating, and recovered hot water will be available to the treatment plant for space heating and for digester heating in linked anaerobic systems.

5.3.10. VSD Process Simplicity and Stability

The biological process was found throughout the testing program to be relatively simple to operate, resistant to upset, and to rapidly recover from disruptions caused by electrical and mechanical system failures. The straightforward process controls consist of providing a supply of food on a relatively uniform basis and providing air. In a full-scale

Table 8
Dual Digestion Using VSD and Mesophilic Anaerobic Digestion

Comparison of anaerobic control with combined system performance

	11 d SRT Anaerobic control	15 d Anaerobic with VERTAD™	11 d Anaerobic with VERTAD™
Solids retention time (d)			
VERTAD™	0	4	4
Anaerobic	11	15	11
Total	11	19	15
Volatile solids reduction (%)			
VERTAD™	0	40	40
Anaerobic	52	49	45
Total	52	70	67
Anaerobic gas production			
Methane (L/d)	2.8	2	2.5
Methane (L/g COD removed)	0.51	0.39	0.36

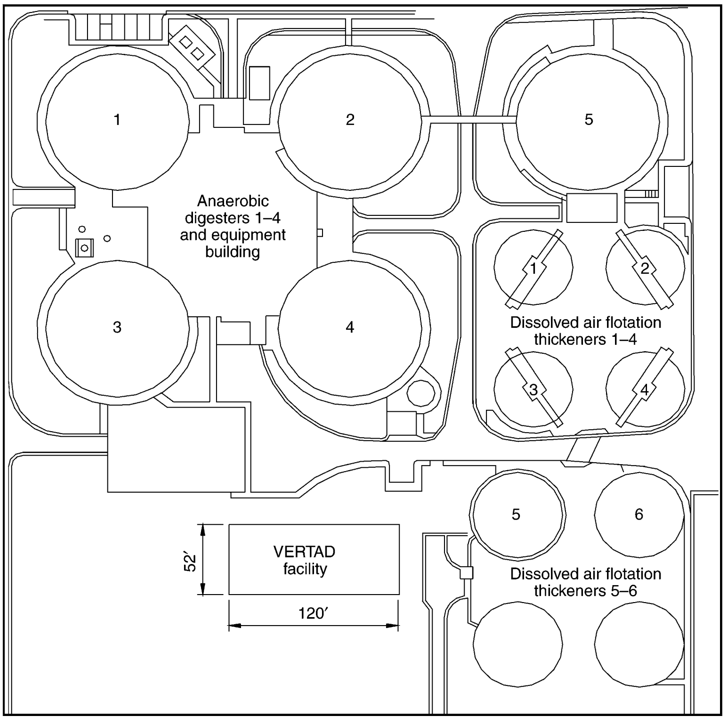
system the operational controls are expected to require less operator attention than in an anaerobic digestion process. The VSD process operates well over a range of pH conditions and temperatures. Although the process does not generate gas, it does produce hot water and does not require the extensive gas handling, cleaning, and safety equipment.

The ability of the process to recover quickly from upset conditions was demonstrated on numerous occasions as the result of power outages and failure at the feed system, boiler, or control system. During these occasions, the process was stressed by lack of food, cooling, and aeration. In all situations the process recovered rapidly.

5.3.11. VSD Followed by Anaerobic Digestion (Dual Digestion)

The process of dual digestion involves the use of an autothermal aerobic digestion process as a pretreatment step before mesophilic anaerobic digestion. In conventional dual digestion systems the aerobic step usually has a contact time of about 24 h and pure oxygen is typically used to support biological metabolism. Dual digestion is a well established class A biosolids process. The VSD process was evaluated as a pretreatment step to mesophilic anaerobic digestion. The impetus to test the combined digestion is the fact that King County treatment facilities, and many biosolids generators, need to maximize solids destruction in order to minimize solids handling costs. The effect of VSD pretreatment on subsequent mesophilic anaerobic digestion was tested using bench scale reactors in studies performed at the University of Washington by Jenny Yoo (3). The results of the dual digestion study are presented in Table 8.

The results indicate that following VSD with mesophilic anaerobic digestion provides additional reduction of volatile solids with the production of significant gas volume. Anaerobic digestion of the VSD product resulted in 67% total volatile destruction with a 4-d SRT in VSD and an 11-d mesophilic anaerobic SRT, and 70% total volatile destruction with a 4-d SRT in VSD and a 15-d mesophilic anaerobic SRT. Comparatively, a control anaerobic digester obtained 52% VS destruction with an 11-d SRT. Although the



Anaerobic digesters are each 7853 ft^2 and includes:
- Capacity = 25 dry t/d
- 1 Anaerobic digester
- Thickening, dewatering, off-gas handling & sludge storage are not included
- Class B biosolids product

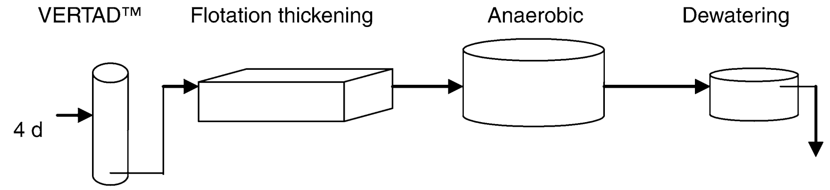
Proposed VERTAD™ facility is 6240 ft^2 and includes:
- Capacity = 25 dry t/d (@ 4-d HRT)
- 2 VERTAD™ digesters
- 2 Head tanks
- 2 Off-gas biofilters
- 6 Heat exchangers
- 2 Flotation tanks
- 1 Sludge storage tank
- 2 Centrifuges
- Class A biosolids product

Fig. 10. Example of VSD retrofit at the King County Sewage Treatment Plant.

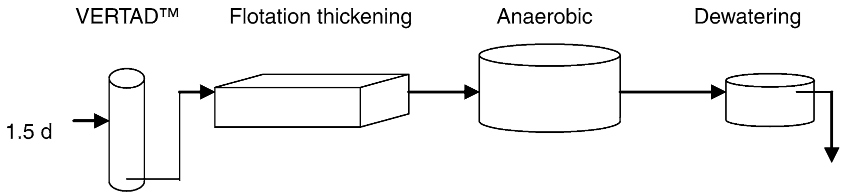
control digester showed more VS reduction in the anaerobic stage than the VSD fed anaerobic digesters (presumably because of the lower VS content in the feed from VSD), the total reduction efficiencies of the dual digestion systems were higher than that of the anaerobic control.

The technologies appear to have a synergy from a performance and operability perspective. For example, the VSD technology readily degrades fats and proteins, whereas anaerobic digestion is capable of cellulose destruction. Observations during the bench scale testing were that the control digester experienced considerable foaming and had mixing problems. The dual digestion systems had no foaming problems and were readily mixed, indicating lower viscosity. This may be attributed to the efficient Org-N and FOG destruction in the VSD process. The ability to float thicken the VSD effluent presents itself as another benefit for the combined system. Thickened product could be fed to anaerobic digestion, allowing operation at higher solids concentrations. The lower volumetric flow associated with the thicker feed would allow for either reduced digester volume requirement or increased solids retention time.

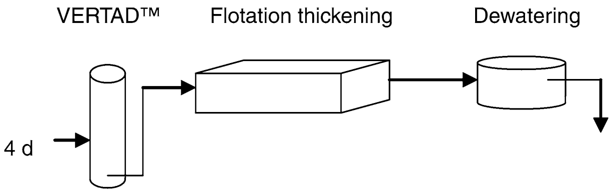
Biosolids with higher total solids concentration would decrease the volumetric flow to dewatering equipment and would likely improve dewatering performance. Several high solids concentration processes are currently being advocated including the anoxic gas flotation process (49). Qualitative indications from the limited dewatering testing of the combined product were that it dewatered to high cake solids (estimated at 24% cake solids) at very low polymer doses (5–6 lb/t).



Concept 1: Class A VERTAD™ digestion followed by anaerobic digestion



Concept 2: Time-temperature VERTAD™ followed by anaerobic digestion



Concept 3: Class A VERTAD™ stand alone digestion process

Fig. 11. VSD (VERTAD™) process flow diagrams.

Incorporation of a post-VSD mesophilic anaerobic digestion step shows considerable promise. The technologies appear to be complementary in many respects from a solids destruction and operability standpoint. This synergy of technologies results in enhanced VS destruction (up to 70%) making VSD an attractive retrofit option for existing anaerobic systems. The minimal footprint requirement for the VSD process makes it an ideal retrofit for facilities that require additional capacity, or current class B biosolids generators that wish to produce class A biosolids. Figure 10 is a schematic showing the STP with a VSD retrofit that could either pretreat the entire sludge stream to class A time-temperature criteria (similar to Concept 2 from Fig. 11) or treat 25 dry t/d to class A biosolids in a stand alone VSD facility (similar to Concept 3 from Fig. 11).

5.3.12. Full-Scale Design and Economics

The results of the demonstration project provided the basis for full-scale design parameters and cost estimates for the VSD process. Planning level designs were developed for three alternatives for solids treatment facilities at a planned future 36 MGD treatment plant in King County. The alternative flow sheets presented in Fig. 11 were developed in detail for the County (3,47). The present worth of capital costs for a system with VSD pretreatment before anaerobic digestion was similar to that of mesophilic anaerobic digestion alone. The present worth of operating cost was significantly

Table 9
Comparison of VSD and Conventional Autothermal Thermophilic Aerobic Digestion

Parameter	ATAD (design)	ATAD (case study)	VERTAD™ (design)
Power usage (kWh/t TS fed)	442	520–641	315
Power usage (kWh/t VS destroyed)	1.52	1.85–2.32	1.27
Aeration ($m^3/h/m^3$ active reactor volume)	4	Not measured	1.7
Time for VS destruction of 40–42% (d)	5–8	12–15	3.5–5
Average system OTE (%)	Not reported	24%	50%

less than conventional anaerobic, primarily because of savings in dewatering and hauling cost in the VSD system.

Additional benefits not accounted for in the capital and operating cost analysis are expected to further improve the comparison, making VSD a favorable choice for King County or a similar community. These additional benefits include, but are not limited to:

a. The value of class A biosolids (the potentially increased market for beneficial reuse).
b. Low grade heat recovered from the process can be utilized for space heating.
c. Decreased land requirements for the VSD process.
d. VSD product synergy with a subsequent anaerobic digestion step (improved mixing, less scum, higher solids concentration, decreased size of dewatering facility, and improved dewatering).
e. Reduction in the NIMBY effect because of minimal odor release, and an aesthetically pleasing (out-of-sight) facility design.

5.3.13. Capital Costs

Except for very small flow facilities, the capital cost of a VSD system is lower than that in conventional plants of similar size. Decreased land requirements, considerably less surface tankage (less concrete), less dewatering equipment and fewer pumps are some of the key elements decreasing the capital cost.

Several factors support the reduced capital costs and land requirements of VSD systems. These factors amount to VSD requiring 10–20% of the total land required for conventional anaerobic plants of equivalent capacity–reducing visual and environmental impact. Some of these factors include:

a. 80% of the bioreactor volume is below grade—eliminating surface tankage.
b. As a result of the high oxygen transfer efficiency in VSD systems, the residence time required in the bioreactor is decreased relative to conventional technologies—making the required reactor volume smaller.
c. The solids are easily float-thickened to 8–12% TS out of the VSD reactor. Float thickening in this manner significantly reduces the size of the downstream dewatering facility.

5.3.14. Operating Costs

The most significant savings realized in both the VSD and the VSB processes relate to the aeration system (3,50). The basis of the VSD process is that the oxygen transfer efficiency is significantly higher than that in a conventional aerobic digestion system because of the pressure at the depth where air is introduced to the bioreactor. In a recent comparison study of the energy requirements between VSD and ATAD processes, it was found that VSD out-performed a conventional ATAD process, operating with 31–45% less energy per pound of VS destroyed in the system. It was also found that the VSD process obtained a doubling in oxygen transfer over the conventional ATAD system with 50% OTE compared with 24% OTE, respectively. These results are summarized in Table 9.

A VSD reactor operating at 4% solids can attain an oxygen transfer efficiency of approx 50%. The resulting aeration power requirement is less than 1.3 kWh/kg (1200 kWh/T) of volatile solids destroyed or 0.35 kWh/kg (315 kWh/T) of total solids treated. Here 1 T = 1000 kg. No additional mixing energy is needed; therefore, the power requirement is lower than the combined aeration and mixing power consumed by conventional aerobic processes.

This air that is economically and efficiently introduced to the bioreactor aids in several other process functions at no incremental cost. Not only does the air satisfy the primary requirement of providing the microbes with dissolved oxygen, it serves as an air lift pump–eliminating the need for mixers in the bioreactor. The air indirectly provides the dissolved gasses necessary for solids flotation in the flotation cell that follows the bioreactor–decreasing the size of the downstream dewatering equipment.

Savings on operating costs have also been realized in the VSD system because of decreased chemical requirements. The VSD biosolids dewater to high cake solids with a very low polymer demand. VSD product can be dewatered to 30–35% solids using a conventional centrifuge, with less than 20 lb/t polymer addition. The exceptionally low polymer consumption reduces operating costs considerably. The ability to effectively dewater biosolids is extremely important because of the high costs associated with hauling and application or landfilling. The high solids content of the dewatered product reduces trucking and disposal costs thus again reducing operating costs. The nutrient value of the class A biosolids product makes it favorable in any beneficial reuse program.

6. CONCLUSIONS

The following conclusions were made based on the results of the demonstration project (3):

a. The VSD (VSD commercially known as VERTAD™) reactor readily circulates THS (4–6% TS); the upper zones are well mixed whereas the lower zone is hydraulically separated, providing strict adherence to the class A pathogen requirements of US EPA 40 CFR 503.
b. The vector attraction reduction and pathogen destruction requirements for class A biosolids were achieved with a 4 d solids retention time (US EPA 40 CFR 503, Alternative 1).
c. Oxygen transfer efficiency was more than 50% when the reactor total solids concentration was at or less than 4.5%.
d. VSD effluent could be easily float thickened to 8–12% TS by pH-shift CO_2 release; thickened product dewatered to more than 30% cake solids with low polymer demand (14 lb/t).
e. Organic nitrogen and fats, oils, and greases were preferentially degraded over organic solids primarily consist of mesophilic anaerobic digestion of VSD effluent provided overall volatile solids destruction of 67% and gas production of 0.36 L CH_4/g COD removed with a combined solids retention time of 15 d (4 d SRT in VSD followed by an 11 d SRT in anaerobic digestion).

f. VSD had low operating cost because of low energy requirements (1.27 kWh/kg VS destroyed), low polymer requirements (14 lb/t), and low trucking/disposal costs (≤30% TS cake).
g. A cost evaluation of full-scale implementation at King County sewage treatment plant indicated that a combined system of VSD and mesophilic anaerobic digestion has a similar present worth of capital and operating costs compared to traditional anaerobic digestion.
h. The VSD process has a minimal footprint requirement making it an ideal retrofit for facilities that require additional capacity, or current class B biosolids generators that wish to produce class A biosolids.

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APPENDIX 1

US Environmental Protection Agency 40CFR503-Appendix B (Pathogen Treatment Processes)

A. Processes to Significantly Reduce Pathogens (PSRP)

1. Aerobic digestion—Sewage sludge is agitated with air or oxygen to maintain aerobic conditions for a specific mean cell residence time at a specific temperature. Values for the mean cell residence time and temperature shall be between 40 days at 20 degrees Celsius and 60 days at 15 degrees Celsius.
2. Air drying—Sewage sludge is dried on sand beds or on paved or unpaved basins. The sewage sludge dries for a minimum of three months. During two of the three months, the ambient average daily temperature is above zero degrees Celsius.
3. Anaerobic digestion—Sewage sludge is treated in the absence of air for a specific mean cell residence time at a specific temperature. Values for the mean cell residence time and temperature shall be between 15 days at 35 to 55 degrees Celsius and 60 days at 20 degrees Celsius.
4. Composting—Using either the within-vessel, static aerated pile, or windrow composting methods, the temperature of the sewage sludge is raised to 40 degrees Celsius or higher and remains at 40 degrees Celsius or higher for five days. For four hours during the five days, the temperature in the compost pile exceeds 55 degrees Celsius.
5. Lime stabilization—Sufficient lime is added to the sewage sludge to raise the pH of the sewage sludge to 12 after two hours of contact.

B. Processes to Further Reduce Pathogens (PFRP)

1. Composting—Using either the within-vessel composting method or the static aerated pile composting method, the temperature of the sewage sludge is maintained at 55 degrees Celsius or higher for three days.

   Using the windrow composting method, the temperature of the sewage sludge is maintained at 55 degrees or higher for 15 days or longer. During the period when the compost is maintained at 55 degrees or higher, there shall be a minimum of five turnings of the windrow.
2. Heat drying—Sewage sludge is dried by direct or indirect contact with hot gases to reduce the moisture content of the sewage sludge to 10 percent or lower. Either the temperature of the sewage sludge particles exceeds 80 degrees Celsius or the wet bulb temperature of the gas in contact with the sewage sludge as the sewage sludge leaves the dryer exceeds 80 degrees Celsius.
3. Heat treatment—Liquid sewage sludge is heated to a temperature of 180 degrees Celsius or higher for 30 minutes.
4. Thermophilic aerobic digestion—Liquid sewage sludge is agitated with air or oxygen to maintain aerobic conditions and the mean cell residence time of the sewage sludge is 10 days at 55 to 60 degrees Celsius.
5. Beta ray irradiation—Sewage sludge is irradiated with beta rays from an accelerator at dosages of at least 1.0 megarad at room temperature (ca. 20 degrees Celsius).
6. Gamma ray irradiation—Sewage sludge is irradiated with gamma rays from certain isotopes, such as \60\ Cobalt and \137\ Cesium, at dosages of at least 1.0 megarad at room temperature (ca. 20 deg. Celsius).
7. Pasteurization—The temperature of the sewage sludge is maintained at 70 degrees Celsius or higher for 30 minutes or longer. [58 FR 9387, Feb. 19, 1993, as amended at 64 FR 42573, Aug. 4, 1999].

APPENDIX 2

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers[a]

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

[a] From ref. 46.

16

Vacuum Filtration

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. FILTRATION DEWATERING SYSTEMS

Filtration can be defined as the removal of solids from a liquid stream by passing the stream through a porous medium, which retains the solids. Figure 1 shows a flow diagram of a filtration system (1).

As indicated on Fig. 1, a pressure drop is required in order for liquid to flowthrough the porous medium (2). This pressure drop can be achieved in four ways that are as follows:

a. By creating a vacuum on one side of the porous medium.
b. By raising the pressure more than the atmospheric pressure on one side of the medium.
c. By creating a centrifugal force on an area of the porous medium.
d. By designing to make use of gravitational force on the medium.

Biosolids filtration dewatering processes use one or more of these driving forces and fall under the general filtration category of surface filters. Surface filters are the type of filtration in which solids are removed and collected in the form of a cake on the upstream side of thin filter media (1).

From: *Handbook of Environmental Engineering, Volume 6: Biosolids Treatment Processes*
Edited by: L. K. Wang et al. © The Humana Press Inc., Totowa, NJ

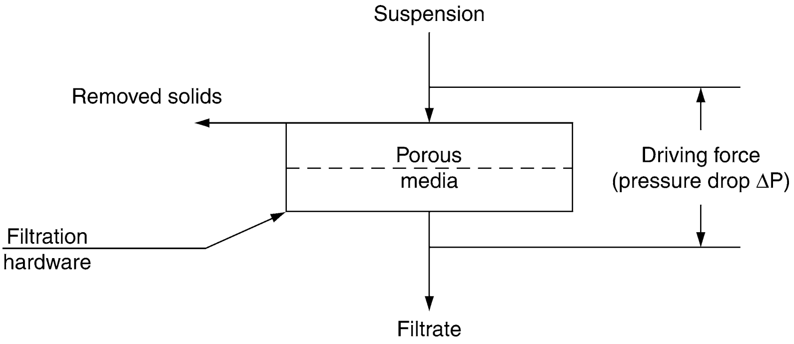


Fig. 1. Flow diagram of a filtration system (*Source:* US EPA).

1.1. Basic Theory

The filtration theory stems from Darcy's original work in the mid-1850s (3). Darcy found that the flow rate of a filtrate through a bed was related to the driving pressure Δp. This relationship is shown in Eq. (1).

$$Q = K\,A\,\Delta p/\mu\,L \tag{1}$$

where Q is the flow rate; K is the bed permeability; A is the bed area; Δp is the driving pressure; μ is the filtrate viscosity; and L is the medium thickness.

Many times, Eq. (1) is written as:

$$Q = A\,\Delta p/\mu\,R \tag{2}$$

where R is the medium resistance = L/K.

Extensive research has been conducted in defining the factors involved and level of influence in dewatering both compressible and incompressible biosolids and other sludges. A comprehensive discussion on filtration has been published by Svarovsky (4). This discussion, through examples, shows the effects of constant pressure filtration; constant rate filtration; constant rate-constant pressure filtration; and variable pressure, and variable rate filtration on both compressible and noncompressible sludges. For persons interested in the detailed mathematical analysis of vacuum filtration, its design equations, and the determination of the equations parameters, the readers are referred to the subject on sludge treatment in the book *Environmental Engineering* by Sincero and Sincero (5).

1.2. Filter Aids

Filter aid is a material such as diatomite, perlite, cellulose, or carbon that serves to improve, or increase the filtration rate by physical means only. Filter aids are not added directly to the biosolids, as a conditioning agent is, but they are added in fixed amount to the porous medium of the particular dewatering equipment. The amount of filter aid added is independent of biosolids concentration. The filter aid literally becomes the "filtering surface" that achieves the liquid/solids separation, and the equipment functions

Table 1
Precoat[a] Process Performance on Fine Particulate Sludges or Biosolids

	Sludge properties			Performance			
Case	Feed solids conc. (%)	Particle size (μm)	Specific resistance $\times 10^7$ (s^2/g)	Solids loading ($lb/ft^2/h$)	Cake solids (%)	Diatomite used (lb/ton dry solids)	Solids capture (%)
1. Mixture alum	0.5	4	354	0.28	26	820	99.9+
WAS[b]-RVPF[c]	5	2	–	1	23	280	99.9+
2. WAS–RVPF	2.2	10	3.2	2.20	25–30	160	99.9+
Conditioned WAS-FP[d]	11.4[e]	–	–	0.30	40–45	–	98.5+
3. WAS–RVPF	1–2	–	40–790	0.55–2.09	26–33	140	99.9+
Conditioned WAS–FP	1–2	–	2–317	0.23–1.44	26–40	200	98
4. WAS–RVPF	1.5	–	53	0.88	29	280	99.9+
Conditioned WAS–RVPF	1.5	–	16.8	2.51	25	120	99.9+
5. Alum RVPF	0.4–0.8	–	–	0.3	25–30	800	99.9+
6. Alum RVPF	8	15	118	1.37	25	120	99.9+

[a] Diatomite.
[b] Waste-activated sludge.
[c] Rotary vacuum precoat filter.
[d] Filter press.
[e] Fly ash conditioning and precoat.
1 $lb/ft^2/h = 4.9\ kg/m^2/h$.
1 t = 0.907 J.
1 lb = 0.454 g.
1 lb/t = 0.5 kg/T.

as a filter holder. In order to perform its function satisfactorily, the filter aid's particles should be inert, insoluble, incompressible and irregularly shaped, porous, and small (6).

Filter aids normally assist in dewatering difficult-to-handle industrial sludges by either vacuum filtration or pressure filtration. In the last 10 yr, research has been performed on the use of filter aids for improved dewatering of municipal wastewater treatment plant biosolids (7). Table 1 lists results obtained from several studies in which diatomite precoated rotary drum vacuum filters were used.

2. PRINCIPLES OF VACUUM FILTRATION

In vacuum filtration, atmospheric pressure, owing to a vacuum applied downstream of the media, is the driving force on the liquid phase that moves it through the porous media. Vacuum filters were patented in England in 1872 by William and James Hart. The first US application of a vacuum filter in dewatering municipal wastewater treatment plant biosolids was in the mid-1920s (8). Until 1960s, the drum or scraper-type rotary vacuum filter was predominant. In 1970s, the belt-type filter with natural or synthetic fiber cloth, woven stainless-steel mesh, or coil springs media has become dominant and widely used

(9,10). Some of the reasons for this wide spread use of vacuum filters for over a period of 60 yr are as follows (1):

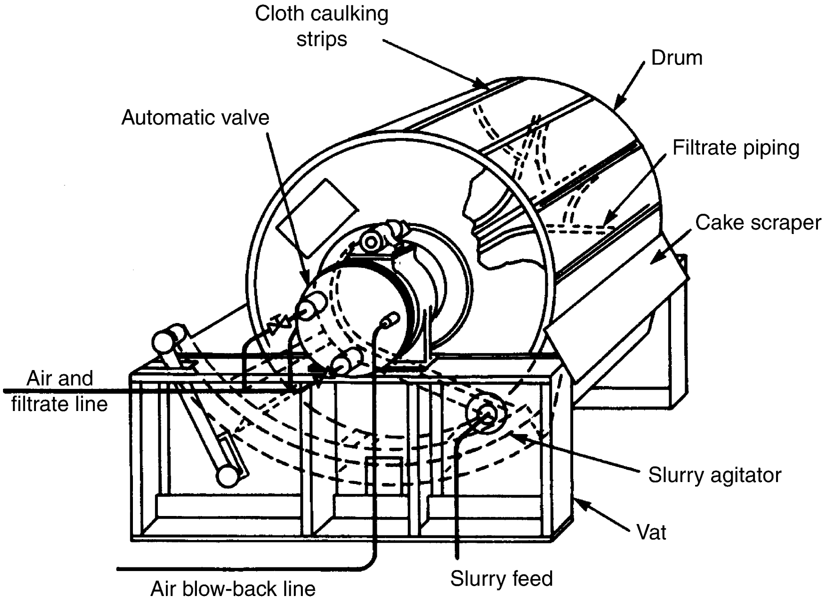
a. Do not require skilled personnel.
b. Have low maintenance requirements for continuous operating equipment.

However, in the late 1980s their use declined and they were replaced by an improved alternative dewatering method using belt filter presses (11–14). Figure 2 shows the cutaway view of a drum or scraper-type, rotary vacuum filter. The unit consists mainly of a horizontal cylindrical drum that rotates, partially submerged, in a vat of conditioned biosolids. The drum surface is divided into sections around its circumference. Each section is sealed from its adjacent section and the ends of the drum. A separate drain line connects each section to a rotary valve at the axis of the drum (15,16). The valve has "blocks" that divide it into zones corresponding to the parts of the filtering cycle. These zones are for cake forming, cake drying, and cake discharging. A vacuum is applied to certain zones of the valve and subsequently to each of the drum sections through the drainlines as they pass through the different zones in the valve. The various operating zones encountered during a complete revolution of the drum are illustrated in Fig. 2.

About 20–35% of the drum surface is submerged in a vat containing the biosolids or sludge slurry. This portion of the drum is referred to as the cake forming zone. Vacuum applied to a submerged drum section causes filtrate to pass through the media and cake to be formed on the media. As the drum rotates, each section is successively carried through the cake forming zone to the cake drying or dewatering zone. This zone is also under vacuum and begins at, and when a drum section carries formed cake out of the biosolids vat. The cake drying zone represents 40–60% of the drum surface and terminates at the point where vacuum is shut off to each successive section. At this point, the biosolids cake and drum section enter the cake discharge zone. In this final zone, cake is removed from the media. Belt-type rotary vacuum filters differ from the drum or scraper-type units, because the drum covering or media-belt leaves the drum. There are essentially two coverings used with belt-type units: coil springs or fiber cloth.

Figure 3 shows a cross-sectional view of a coil spring belt-type rotary vacuum filter. This filter uses two layers of stainless steel coils arranged around the drum. After the dewatering cycle, the two layers of springs leave the drum and are separated from each other. In this way, the cake is lifted off the lower layer of springs and can be discharged from the upper layer. Cake release is essentially never a problem. After cake discharge, the coils are washed and returned to the drum. The coil filter has been widely used for most types of sludge (17–19).

However, biosolids with particles that are both extremely fine and resistant to flocculation dewater poorly on coil filters (1). The coil springs, which have 7–14% open area, act to support the initial solids deposit, which in turn serves as the filtration medium. Because of the open area of the springs, it is important that the feed solids concentration be high; that is, it should contain sufficient fibrous material to prevent the loss of fine solids. Sludges with particles that are both extremely fine and resistant to flocculation dewater poorly on coil filters, and solids capture is low. A cloth medium is required when filtering unthickened biosolids that are predominantly secondary solids.



Cutaway view of a drum or scraper-type rotary vacuum filter

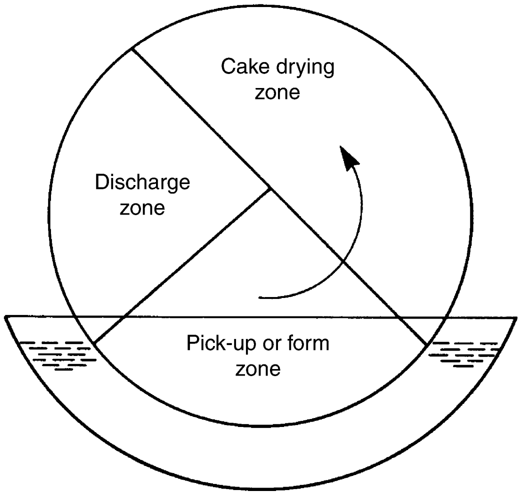


Fig. 2. Cutaway view and operating zones of a drum rotary vacuum filter (*Source:* US EPA).

Figure 4 shows a schematic cross-section of a fiber cloth-belt and rotary vacuum filter. Media on this type unit leaves the drum surface at the end of the drying zone and passes over a small-diameter discharge roll to facilitate cake discharge. Washing of the media occurs after discharge and before it returns to the drum for another cycle. Normally, this type of filter has a small-diameter curved bar between the point

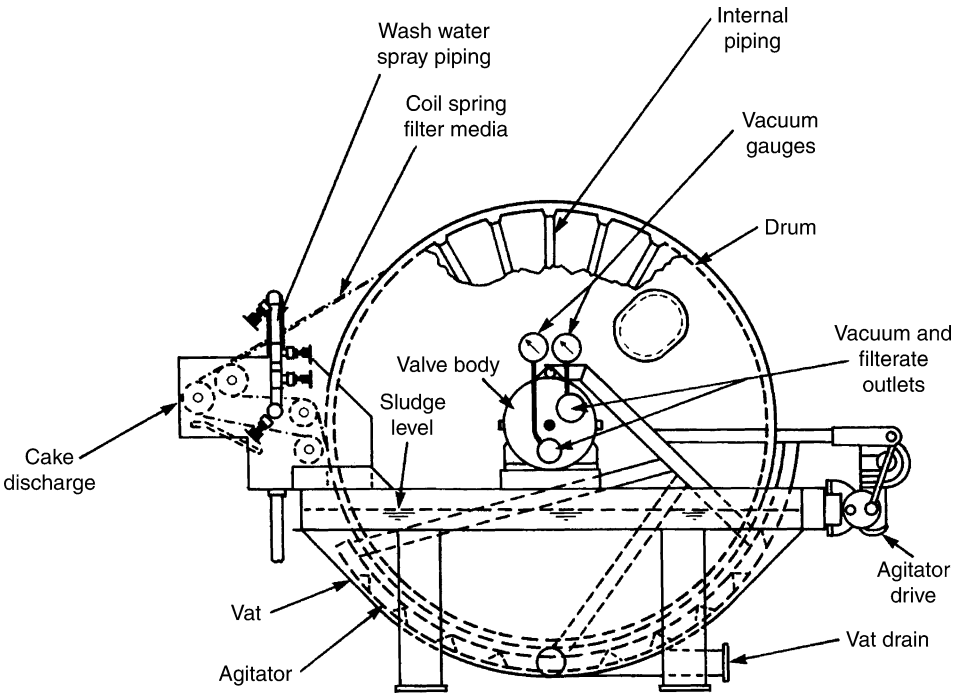


Fig. 3. Cross-sectional view of a coil spring-belt type-rotary vacuum filter (*Source:* US EPA).

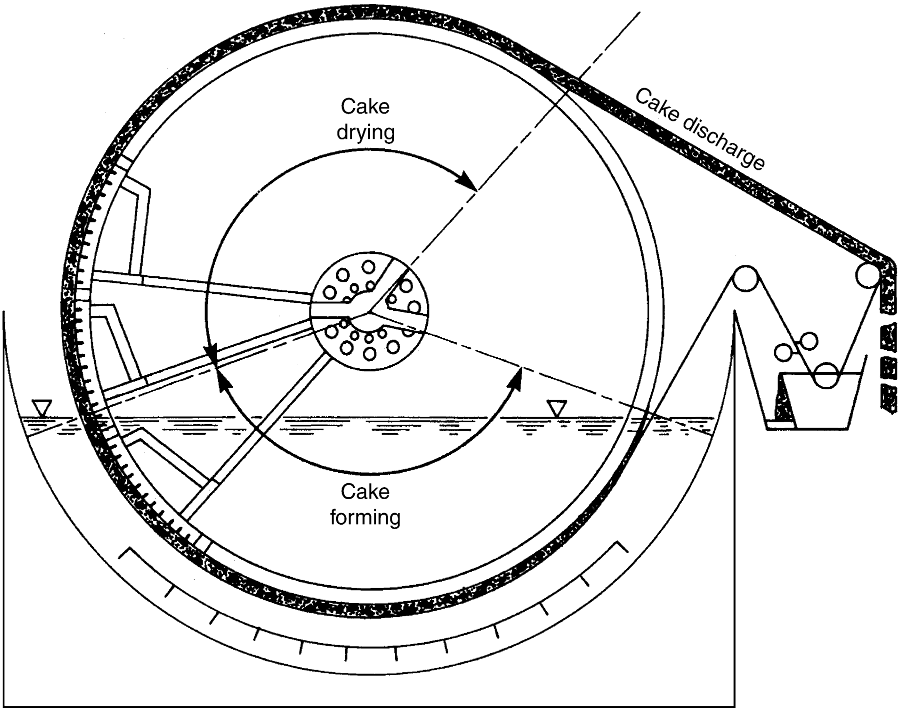


Fig. 4. Cross-sectional view of a fiber cloth-belt type-rotary vacuum filter (*Source:* US EPA).

where the belt leaves the drum and the discharge roll. This bar aids in maintaining belt dimensional stability. In practice, it is frequently used to ensure adequate cake discharge. Remedial measures, such as addition of scraper blades, use of excess chemical conditioner, or addition of fly ash, are sometimes required to obtain cake release from the cloth media (1). This is particularly true at wastewater treatment plants, which produce biosolids that are greasy, sticky, and/or contain a large quantity of activated sludge.

A summary of the common major equipment variations are:

a. **Scraper-discharge mechanism.** The filter drum operates continuously with a vacuum pickup forming and filtering zone, a vacuum cake drying zone, and a pressure blow back or discharge zone. A positive air pressure is maintained in the segment just ahead of the biosolids scraper blade to aid in removal of the dried cake. A fine spray may be used to clean the filter medium with a catching through beneath to dispose of the washings.
b. **String discharge filters.** Closely spaced strings around the filter drum, the medium, and a set of discharge and return rolls carry the biosolids cake and then free it from the medium and discharge it to a hopper. The strings pass through a set of aligning combs before returning to the drum.
c. **Belt-medium filters.** A traveling woven cloth or metal belt serves as the filter medium and transports the biosolids cake to the discharge roll in a manner similar to that of the string discharge filters. If desired, the belt can be washed on both sides, before positioning back on the drum.
d. **Coil-medium filters.** Two layers of stainless steel springs wrapped around the drum act as the filter medium. When the two layers of springs leave the drum they separate in such a manner that the biosolids cake is lifted off the lower layer of coil springs and discharged off the upper layer with the aid of a positioned tine bar. The two coil spring layers are then washed separately by spray nozzles and returned to the drum.

3. APPLICATION AND PERFORMANCE

Vacuum filters have probably been used to dewater more types of municipal wastewater treatment plant sludges than most other mechanical dewatering equipment. Since the mid-1920s, more than 1700 vacuum filters have been installed in over 800 United States municipalities (1). The era of vacuum filtration is now history. Improvements in other dewatering devices, as well as the development of new dewatering devices, have permitted municipalities to dewater their biosolids and other sludges as well as they could with vacuum filters, but at lower operation and maintenance costs (20,21).

As with all types of mechanical dewatering equipment, optimum performance is dependent upon the type of sludge and its solids concentration, type and quality of conditioning, and how the filter is operated. Selection of vacuum level, degree of drum submergence, type of media, and cycle time are all critical to optimum performance. Tables 2 and 3 contain performance data for cloth and coil media rotary vacuum filters for the sludge types indicated. Tables 4 and 5 contain specific operating data for several wastewater treatment plants using cloth media and coil media (1).

The efficiency of solids removal, or percent solids recovery, is the actual percentage of feed solids recovered in the filter cake. Solids removal on vacuum filters with adequate chemical conditioning, range from about 85% for coarse mesh media to 98% with close weave, long nap media. The recycled filtrate solids impose a load on the treatment plant and should normally be kept to a practical minimum. However, it may be necessary to

Table 2
Dewatering Performance Data for Rotary Vacuum Filters-Cloth Media

Types of sludge	Feed solids conc. (%)	Chemical dosage[a] (lb/t dry solids) FeCl$_3$	CaO	Yield[b] (lb dry solids/ft^2/h)	Cake (solids %)
Raw primary (P)	4.5–9	40–80	160–200	3.5–8	27–35
WAS	2.5–4.5	120–200	240–360	1–3	13–20
P + WAS	3–7	50–80	180–240	2.5–6	18–25
P + trickling filter (TF)	4–8	40–80	180–240	3–7	23–30
Anaerobically digested					
P	4–8	60–100	200–260	3–7	25–32
P + WAS	3–7	80–120	300–400	2–5	18–25
P + TF	5–10	80–120	250–350	3.5–8	20–27
Aerobically digested no primary clarification	2.5–6	60–140	150–240	1.5–4	16–23
Elutriated anaerobic digested					
P	5–10	50–80	0–100	4–8	27–35
P + WAS	4.5–8	60–120	0–150	3–6	18–25
Thermally conditioned P + WAS	6–15	0	0	4–8	35–45

[a] All values shown are for pure FeCl$_3$ and CaO. They must be adjusted for anything else.

[b] Filter yields depend to some extent on feed solids concentrations, increasing the concentration normally gives a higher yield.

1 lb/t = 0.5 kg/T.

1 lb/ft^2/h = 4.9 kg/m^2/h.

Table 3
Dewatering Performance Data for Rotary Vacuum Filters-Coil Media

Types of sludge	Feed solids conc. (%)	Chemical dosage[a] (lb/t dry solids) FeCl$_3$	CaO	Yield[b] (lb dry solids/ft^2/h)	Cake (solids %)
P	8–10	40–80	160–240	6.5–8	28–32
TF	4–6	40–60	100–140	6–8	20–28
P + WAS	3–5	20–60	180–220	2.5–4	23–27
Anaerobically digested					
P + TF	5–8	50–80	240–320	4–6	27–33
P + WAS	4–6	50–80	200–300	3.5–4.5	20–25
Elutriated anaerobically digested primary	8–10	20–50	30–120	4–8	28–32

[a] All values shown are for pure FeCl$_3$ and CaO. This must be adjusted for anything else.

[b] Filter yields depend to some extent on feed solids concentration. Increasing the solids concentration normally gives a higher yield.

1 lb/t = 0.5 kg/T.

1 lb/ft^2/h = 4.9 kg/m^2/h.

Table 4
Specific Operating Results of Rotary Vacuum Filters-Cloth Media

			Conditioner used percent by weight[b]				
Location	Sludge type[a]	Feed solids conc. (%)	FeCl$_3$	CaO	Cake (Solids %)	Yield (lb dry solids/ft^2/h)	Filtrate (mg/L)
Willoughby Eastlake, OH	P + WAS + septic	4–6	3	14	20	2.8–4.8	–
Tamaqua, PA	Anaerobically digested (P + WAS)	6	3	26	18	3	SS 20–30
Grand rapids, MI	Thermally conditioned (P + WAS)	10–15	None	None	50	6	SS 5000 BOD$_5$ 10,000
Fort Atkinson, WI	WAS	3–4	6	16	19	3–3.5	–
Frankenmuth, MI	WAS	3.7	8	14	15	3.2	–
Oconomowoc, WI	Anaerobically digested (P + WAS)	2.3	6	20	18	2.5–3	SS 500–1100 BOD$_5$ 10
Genessee City, MI	P + WAS	8	–	16	27	5.6	–

[a] WAS, waste-activated sludge; P, primary sludge.
[b] Numbers shown are based on pure FeCl$_3$, and pure CaO.
1 lb/ft^2/h = 4.9 kg/m^2/h.

Table 5
Specific Operating Results of Rotary Vacuum Filters-Coil Media

		Conditioner used percent by weight[b]			
Location	Sludge type[a]	FeCl$_3$	CaO	Cake (Solids %)	Yield (lb dry solids/ft^2/h)
Blytheville, AR	TF	36	94	33.1	10.4
York, PA	Anaerobically digested (P + WAS)	80	250	21.1	4.7
Wyomissing Valley, PA	Anaerobically digested TF	62	272	18.2	6
Bayonne, NJ	Anaerobically digested P	28	62	30.9	7.8
Woodbridge, NJ	P	40	240	29.7	8
Shadyside, OH	Anaerobically digested (P + WAS)	64	310	29	4.2
Arlington, TX	TF	64	174	25.2	8.8

[a] WAS, waste-activated sludge; P, primary sludge. No data available for feed solids and filtrate.
[b] Numbers shown are based on pure FeCl$_3$, and pure CaO.
1 lb/ft^2/h = 4.9 kg/m^2/h.

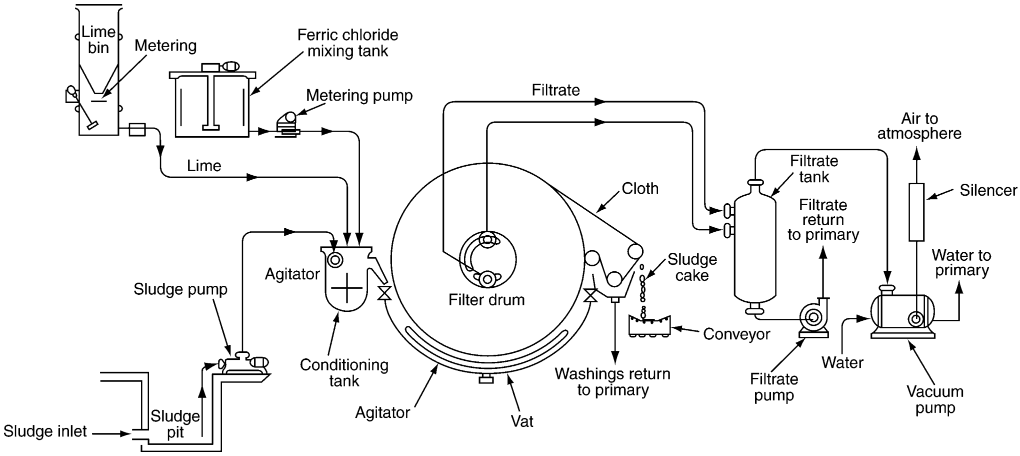


Fig. 5. Rotary vacuum filter system (*Source:* US EPA).

reduce the percent recovery in order to deliver more filter output and thus keep up with biosolids production.

4. AUXILIARY EQUIPMENT

Normally, rotary vacuum filters are supplied with auxiliary equipment including vacuum pump, filtrate receiver and pump, and biosolids conditioning apparatus. Figure 5 shows a typical complete rotary vacuum filter process. Usually, one vacuum pump is provided for each vacuum filter, although some larger plants use less than one pump per filter and the pumps connect to a common header. Until the 1960s, reciprocating type dry vacuum pumps were generally specified, but since the early 1970s wet type vacuum pumps were universally used. The wet type pumps are more easily maintained and provide sufficient vacuum. Wet type pumps utilize seal water and it is essential that satisfactory water be used. If the water is hard and unstable, it may be necessary to prevent carbonate buildup on the seals through the use of a sequestering agent. Normally, the vacuum pump requirements are 1.5–2 adiabatic ft^3/min of air/ft^2 of drum surface area at 20 in. of mercury vacuum (1.5 m^3/min/m^2 at 69 kN/m^2). This is true unless the expected yield is more than 40–50 lb/ft^2/h (20–25 kg/m^2/h) and extensive biosolids cake cracking is expected. In the latter case, an air flow 2.5 times higher should be used (1).

Each vacuum filter must be supplied with a vacuum receiver located between the filter valve and the vacuum pump. The principal purpose of the receiver is to separate the air from the liquid. Each receiver can be equipped with a vacuum-limiting device to admit air flow if the design vacuum is exceeded (a condition that could cause the vacuum pump to overload). The receiver also functions as a reservoir for the filtrate pump suction. The filtrate pump must be sized to carry away the water separated in the vacuum receiver, and normally it is sized to provide a capacity two to four times the design biosolids feed rate to the filter (1).

The filtrate pump should be able to pump against a minimum total dynamic head of between 40 and 50 ft (12–15 m), which includes a minimum suction head of 25 ft (7.5 m).

Centrifugal-type pumps are commonly used but can become air bound unless they have a balance or equalizing line connecting the high point of the receiver to the pump. Typically, nonclogging centrifugal style pumps are used with coil filters because they permit somewhat higher solids concentration in the filtrate. Self-priming centrifugal pumps are used most frequently, as they are relatively maintenance free. Check valves on the discharge side of the pumps are usually provided to minimize air leakage through the filtrate pump and receiver to the vacuum pump.

5. OPERATING FACTORS

5.1. Machine Variables

The principle machine variables that impact vacuum filter operation are as follows:

a. Filter media used.
b. Quantity of wash water used.
c. Drum speed.
d. Vacuum level.
e. Conditioning chemicals—type and dosage.
f. Drum submergence.
g. Vat agitation.

Establishment of the drum speed, optimum vacuum level, conditioning methods, drum submergence and optimum media selection can all be accomplished on bench scale. The drum speed establishes the cycle time and the submergence sets the form time and drying time. Normally, the media selection is made at the time of equipment start-up by the equipment supplier. The trend over the years has been to select a monofilament fabric, because they seem the most resistant to blinding and have a reasonably long life.

A change in conditioning procedures, biosolids mixture, or biosolids holding time (time held before conditioning and dewatering) impact on the efficiency of a given medium. With belt filters, wash water at a pressure of at least 480 kN/m^2 (70 psi) must be available. Throughput is usually estimated from data gathered with clean media. Generally, it is observed that where there is insufficient cloth washing, increasing the amount of wash water will increase the machine throughput and will help to increase cake dryness. Vat agitation is necessary for proper cake formation, but over- agitation will result in breaking up the biosolids floc, poor solids capture, and lower feed rates. The addition of scraper blades, use of excess chemical conditioner, or addition of fly ash are sometimes required to obtain cake release from cloth media vacuum filters.

5.2. Filter Media

A major process variable is the filter media. The ideal media performs the desired liquid/solid separation and gives a filtrate of acceptable clarity (22). Further, the filter cake discharges readily from it, and it is mechanically strong enough to give a long life. The media must be chemically resistant to the materials being handled and provide minimal resistance to filtrate flow. A further characteristic to be minimized is “blinding” or clogging. All the characteristics aforementioned need to be evaluated during the selection procedure. One must, therefore, through experience, or bench or pilot-scale rotary

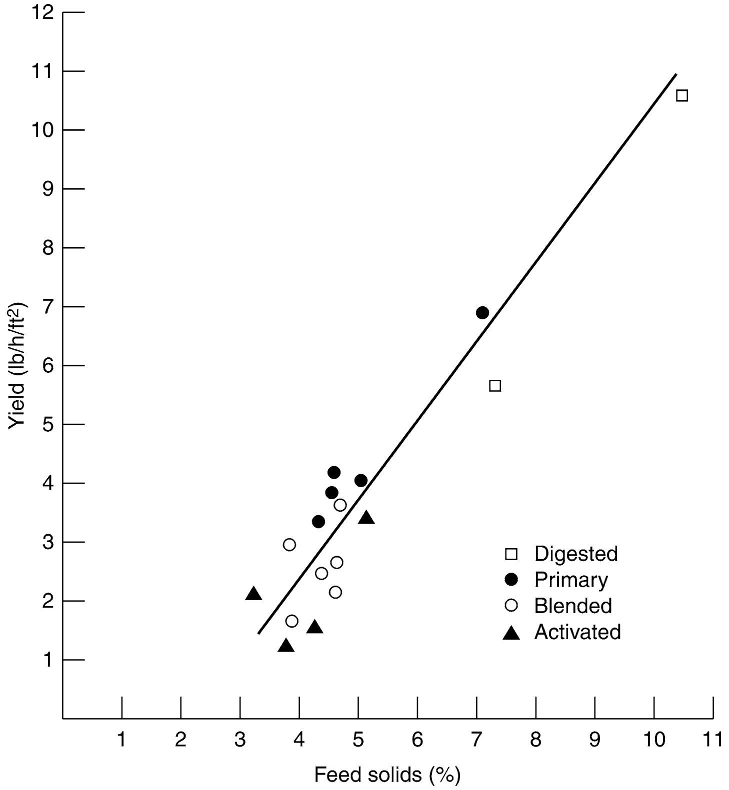


Fig. 6. Rotary vacuum filter productivity as a function of feed sludge suspended solids concentration (*Source:* US EPA).

vacuum filter testing, select the best media in terms of porosity, type of weave, material of construction, and so on, for a particular sludge. Normally, this selection is made at the time of equipment start-up by the equipment supplier (23,24).

5.3. *Solids Feed Content*

The higher the feed suspended solids concentration of the biosolids, the more will be the yield or production rate of the rotary vacuum filter (Fig. 6) and the cake suspended solids concentration (Fig. 7). Generally, municipal wastewater treatment plant biosolids are not concentrated beyond about 10% solids, because above this concentration, the biosolids become difficult to pump, mix with chemicals, and to distribute after conditioning to the filter. In addition, to increased production rates, higher biosolids feed concentrations result in lower chemical dosage rates and lower cake moistures (1,5,13,25,26). Both of these consequences affect the cost of biosolids dewatering and ultimate disposal.

Generally, the lowest biosolids feed suspended solids concentration for successful vacuum filtration is considered to be 3%. Less than this concentration it becomes difficult to produce filter cakes thick enough or dry enough for adequate discharge. For this

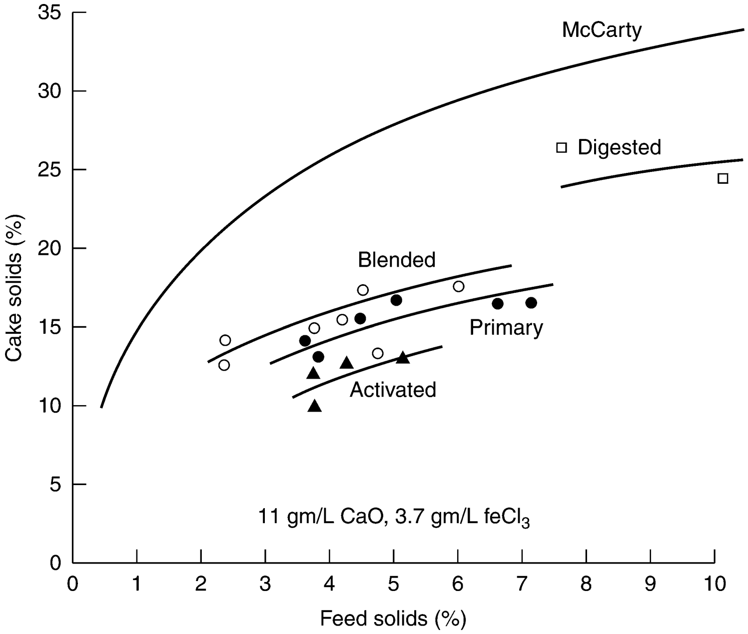


Fig. 7. Sludge cake total solids as a function of the feed sludge suspended solids concentration (*Source:* US EPA).

reason, it is extremely important that the design and operation of the preceding biosolids processes take into consideration the need for an optimal solids concentration when dewatering on vacuum filters.

6. PHYSICAL AND PROCESS CONTROL

6.1. Physical Control

Filter physical control is accomplished by drum speed, vacuum, and biosolids feed rate. The drum speed is controlled by a variable speed drive.

The vacuum is controlled by (27):

a. **Amount of conditioning.** Proper conditioning causes the biosolids to release their water allowing the cake to open up and lowering the vacuum requirements.
b. **Drum speed.** The slowest drum speed produces the thickest, driest, cake, and the lowest vacuum. As the drum speeds up it has less time to remove the water and the vacuum rises with the drum speed.
c. **Biosolids level in the filter vat.** A full vat provides maximum contact time and minimum drying time, resulting in a thicker cake and the highest vacuum. As the vat level is lowered the vacuum drops.
d. **Mechanical devices.** Some systems might be equipped with a spring loaded vacuum release valve, which can be set to open at any desired vacuum level.

6.2. Process Control

Control of the vacuum filter systems should be based on performance. The performance of vacuum filters might be measured by various criteria such as the yield, the efficiency of solids removal, and the cake characteristics. Each of these criteria is of importance, but one or the other might be particularly significant in a given wastewater treatment plant (27).

a. **Yield.** Yield is the most common measure of filter performance. The yield is the filter output and is expressed in terms of pounds of dry total solids in the cake discharged from the filter, per square foot of effective filter area per hour ($lb/ft^2/h$) or $kg/m^2/h$.
b. **Efficiency.** The second measure of filter performance is the efficiency of solids removal. Basically, the vacuum filter is a device used for separating solid matter from liquid, and the actual efficiency of the process is the percentage of feed solids recovered in the filter cake. Solids removal on vacuum filters range from about 85% for coarse mesh media to 99% with close weave, long nap media. The recycled filtrate solids impose a load on the plant treatment units, and normally should be kept to a practical minimum. However, it may be necessary to reduce the filter efficiency in order to deliver more filter output and thus keep up with biosolids production.
c. **Filter cake quality.** The filter cake quality is another measure of filter performance, depending upon cake moisture and heat value. Cake solids content varies from 20 to 40% by weight, depending upon the type of biosolids handled and the filter cycle time and submergence. Delivery of a very dry cake does not necessarily indicate good filter performance. Cake moisture should be adjusted to the method of final disposal; it is inefficient to dry the cake more than is required. When the dewatered biosolids are incinerated, raw biosolids cake having fairly high moisture content can be burned without auxiliary fuel because of the higher volatile content, while digested biosolids cake will have to be drier to burn without make-up heat.
d. **Chemical conditioning.** If chemical conditioning is used, the operator should determine the best conditioning chemical for the feed biosolids. If the character of the feed biosolids is subject to change, an evaluation of conditioning agents should be made after each change. Once an effective conditioner has been selected, the next task is to determine the best chemical dosage rate. One or more of the variables should be held constant and the others varied systematically to develop a series of conditioning performance curves. The best chemical conditioning considering cost and required performance can then be determined.
e. **Tank agitation.** Because all biosolids vary, determine the best procedure for operation of the filter vat agitator by experience. Some biosolids might require continuous use of the agitator; whereas others might be best with no vat agitation (in this case the biosolids must be without agitation from start-up).
f. **Heat-treated biosolids.** The effect of heat treatment on various municipal biosolids is to make all types of biosolids readily dewaterable by vacuum filtration with minimum chemical conditioning. P sludges have been dewatered at rates as high as 40 $lb/ft^2/h$ and waste activated biosolids at 7 $lb/ft^2/h$. Mixtures of P and secondary biosolids subjected to heat treatment should produce yields well over 10 $lb/ft^2/h$.
g. **Optimum operation.** The filter can be operated for maximum biosolids cake output, for the lowest chemical cost, for the driest cake or any combination of these. All that is necessary is to strike a balance between all the controls for the desired output. Once a balance is achieved, it should be easy to maintain by making small adjustments. Large changes in any one of the operating parameters will affect all the others which mean striking a new balance.
h. **Cake drying.** The biosolids cake should not crack until just before it drops off the fabric. This will result in the vacuum continuously pulling air through the biosolids, drying it until the last possible moment rather than just pulling air through the cracks.
i. **Production.** In general, the filter produces more cake as it runs faster; however, as it is hard to judge production between a thin, fast moving cake, and a thick, dry cake, the

production should be based on the biosolids pump speed. The higher biosolids pumping rate corresponds to a higher production. The optimum filter drum speed is the fastest speed that will produce a clean discharge of the cake. An exception to this may occur when dewatered biosolids are to be incinerated and a very dry cake is desired. In this case, moisture content and incinerator capacity govern the drum speed.

j. **Inspection.** The quality of the cake should be observed along with the breakup of the cake as it falls from the filter fabric. After gaining some operating experience it should be possible to roughly judge the operation of the filter by visual appearance.
k. **Odor.** Odor is generated by the vacuum filtration process, but proper preconditioning, chemical conditioning, and ventilation should minimize the problem.
l. **Sampling and analysis.** Samples might be obtained through valves provided in the respective system piping. If sampling points are not provided, they should be installed to facilitate operation and control of the process. Samples of the supernatant can be obtained at the overflow weir. Samples should be analyzed according to procedures specified in *Standard Methods* (28).

7. UPGRADING EXISTING UNITS

As a result of high power costs and the heavy use of inorganic conditioners, vacuum filters are not often selected for use in new facilities. In some cases, refurbishing old equipment may be indicated to minimize capital costs. However, generally older units should not be used for standby capacity unless they are refurbished. The material discussed in this chapter is included to provide guidance primarily for such applications.

The following lists some of the advantages and disadvantages of vacuum filtration relative to other dewatering processes (11):

1. Advantages.
    a. Operation is easy to understand because formation and discharge of biosolids cake are easily visible.
    b. Does not require highly-skilled operator.
    c. Will continue to operate even if the chemical conditioning dosage is not optimized, although this may cause discharge problems.
    d. Coil spring medium has very long life compared with any cloth medium.
    e. Has a low maintenance requirement for a continuously operating piece of equipment, except in certain cases with lime conditioning.
2. Disadvantages.
    a. Consumes a large amount of energy per unit of biosolids dewatered.
    b. Vacuum pumps are noisy.
    c. Lime and ferric chloride conditioning can cause considerable maintenance and cleaning problems.
    d. The use of lime for conditioning can produce strong ammonia odor with digested biosolids.
    e. Best performance is usually achieved at feed solids of 3–4%.
    f. Ferric chloride and lime conditioning costs are higher than polymer conditioning costs. Polymer conditioning is not always effective on vacuum filters.

The significant points to be examined, if refurbishing or reusing vacuum filters, are as follows:

a. Media selection.
b. Feed solids.

c. Conditioning requirements and the design of the conditioning subsystem.
d. Biosolids holding time before and after conditioning.
e. Filtration rate.

Equipment sizing might be accomplished by using information from Tables 2 through 5. These data may be augmented through the use of the filter leaf test. A series of leaf tests will provide a range of values for solids loading and cake solids. The design conditions may be selected from these findings. These procedures have been developed over many years and have excellent reproducibility and a high degree of confirmation.

8. CASE HISTORY

This section is adapted from a US EPA-sponsored investigation (29). Figure 8 shows the 1977 flow diagram for the 13-MGD (34 m^3/s) Lakewood, Ohio, wastewater treatment plant. The biosolids being handled at this plant have changed several times since the facility was built in 1938. At that time, the plant was designed for primary treatment, with biosolids being anaerobically digested and dewatered on sand drying beds. Secondary treatment was added in 1966. Gravity thickeners, two new anaerobic digesters, two vacuum filters, and a flash dryer were installed to handle additional biosolids. In 1974 and 1975, the plant was further upgraded. Alum (aluminum sulfate) was added to the aeration basin effluent channel for phosphorus removal, and the biosolids handling system (filters and dryer) operating schedule was extended to two shifts. Finally, in 1977, the plant was returned to single shift biosolids handling, and excess liquid biosolids were hauled to land disposal.

The Lakewood plant has two polyethylene cloth belt rotary vacuum filters. Only one can be operated at a time because of the limited capacity of the flash dryer. Each filter has an effective area of 376 ft^2 (35 m^2) and operates best at a drum speed of one revolution per eight minutes and a drum submergence between 30 and 36 in. (0.76–0.91 m). A filter is operated 5 d/wk in either one or two 6.5-h shifts/d. Conditioning chemical dosages are approx 275 lb of dry lime (pebble lime-72% CaO) per ton of dry feed solids (137 kg/T) and 30 lb of $FeCl_3$ (liquid at 40% $FeCl_3$) per ton of dry feed solids (15 kg/T).

Before 1975, before alum was added for phosphorus removal (63 mg/L alum added), the average total solids concentration of the digested biosolids (vacuum filter solids feed) was 4.45%. On the average, the biosolids were dewatered to 23.8% solids. After alum addition, the feed solids concentration increased to 6.5%, but the dewatered cake percent dropped to 21.4%.

Table 6 indicates operational costs for 6.5 h and 13 h/d operations based on before and after alum addition for phosphorus removal. Because of the increase in the number of tons from 650 dry t/yr (590 T/yr) in 1974 to 1820 dry t/yr (1651 T/yr) in 1976, the treatment cost per ton of dry total solids was not relatively more than it was in 1974.

9. COSTS

Economic aspects of biosolids dewatering are analyzed in detail by US EPA Report (30) and Bennett et al. (31). Figure 9 gives the 1975 capital cost as a function of filter area for rotary vacuum filters (1). As an example, a 400 ft^2 (37.2 m^2) area filter would cost 400,000 USD. As this number is based on a June 1975 cost, it must be adjusted to the current year design period. Using the Cost Index for Utilities (Appendix A), the cost

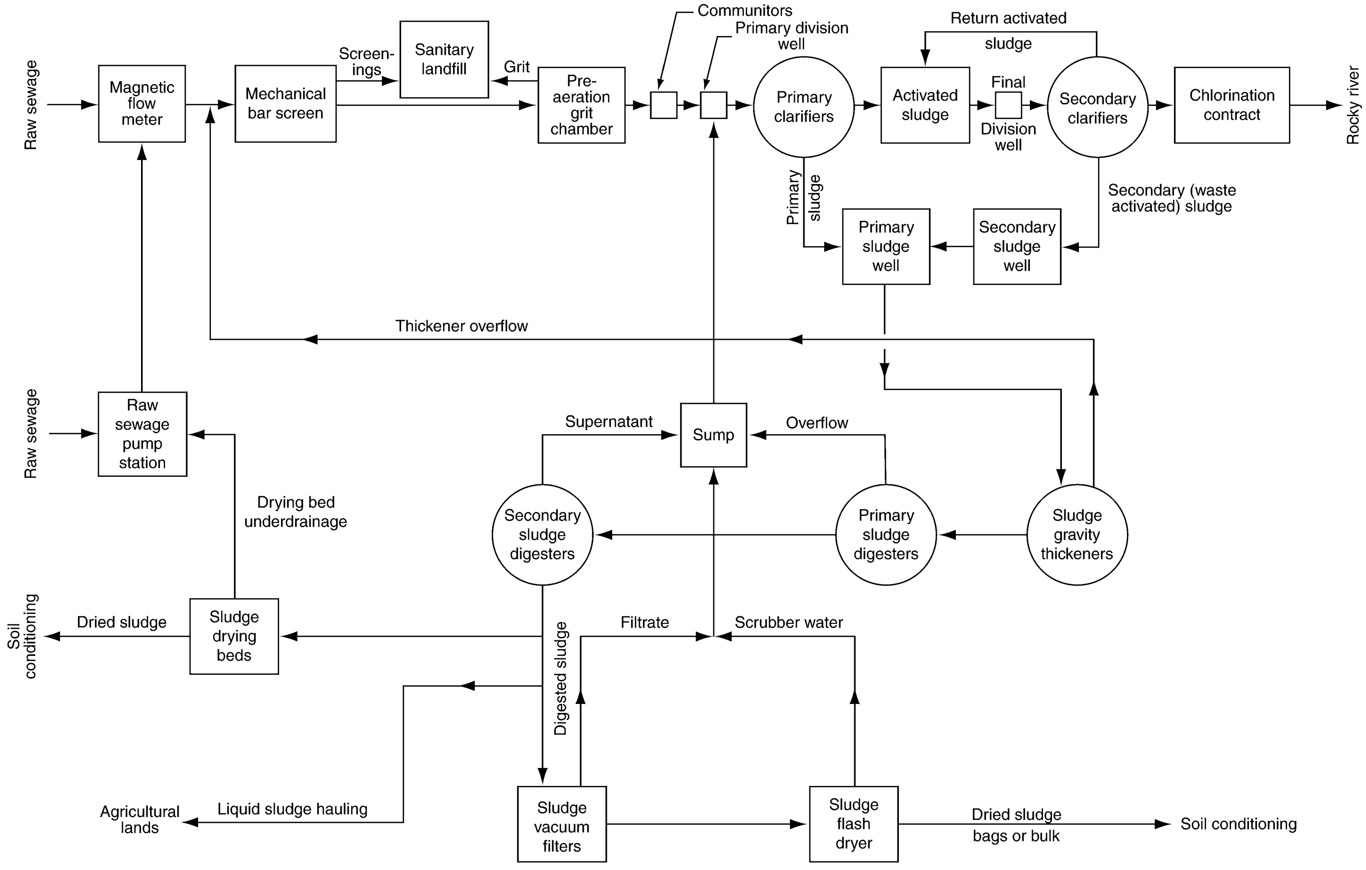


Fig. 8. Lakewood, OH wastewater treatment plant flow diagram (*Source:* US EPA).

Table 6
Operational Cost of Lakewood, OH Vacuum Filter Operations

Item	Single shift operation in 1974 (2006 USD/t dry solids)	Double shift operation in 1976 (2006 USD/t dry solids)
Ferric chloride and lime	24.69	24.69
Electricity	5.50	3.58
Maintenance supplies	3.08	3.06
Maintenance and repair labor	10.12	9.99
Operational labor	9.60	17.35
Overhead	6.25	8.63
Total	59.24	67.30

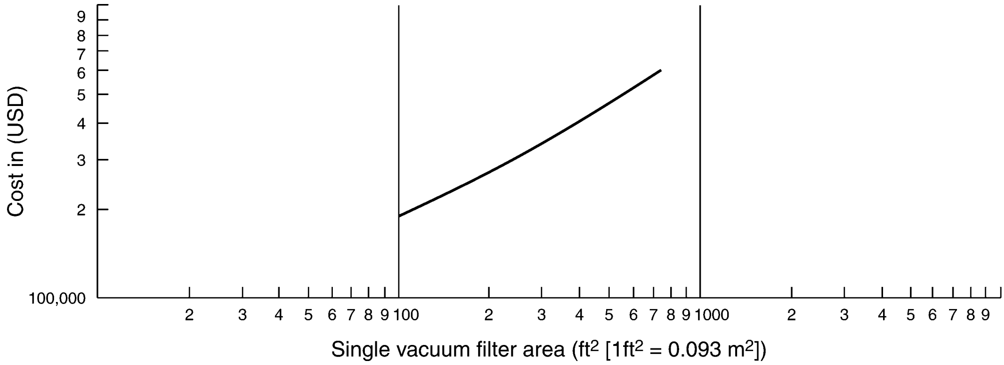


Fig. 9. 1975 Capital cost for rotary drum vacuum filters (*Source:* US EPA).

would be multiplied by a factor of 528.12/190.49 = 2.77 (32). Hence, the cost for the 400 ft^2 (37.2 m^2) area filter would be 1,108,000 USD in 2006 USD. Costs include those for filter, auxiliary equipment, piping, and building.

The labor requirements indicated in Fig. 10 are given as a function of average area in use and include: start-up time and clean-up after the filter run, operation of filter, and operation of biosolids pumping and conditioning facilities before the treatment. As an example, a vacuum filter having 400 ft^2 (37.2 m^2) of filter area would require 550 man-hours of operation and maintenance per year and would be included in cost analysis.

Figure 11 gives the power consumption as a function of filter area. As an example, a vacuum filtration area of 400 ft^2 (37.2 m^2) would require 330,000 kWh/yr (1200 GJ/yr) of electrical energy. If power costs are 0.08 USD/kWh, the cost would be 52,800 USD annually. Energy requirements for solids dewatering is estimated to be 7% (33) of the total plant energy usage (*see* Table 7). Operating parameters used were based on 2 adiabatic ft^3 of air/min/ft^2 (10 L/s/m^2), 20 in. of vacuum (68 KN/m^2), and a total dynamic head of 50 ft (15 m) for the filtrate pump. Power required includes that for drum drive, discharge roller, and vat agitator, but does not include other accessory items, such as biosolids feed pump or chemical feed system.

Figure 12 shows a curve developed for estimating rotary drum vacuum filter maintenance material cost as a function of filter area. As an example, for a filtration area of 400 ft^2 (37.2 m^2), a designer would estimate a yearly materials cost of 4000 USD. As

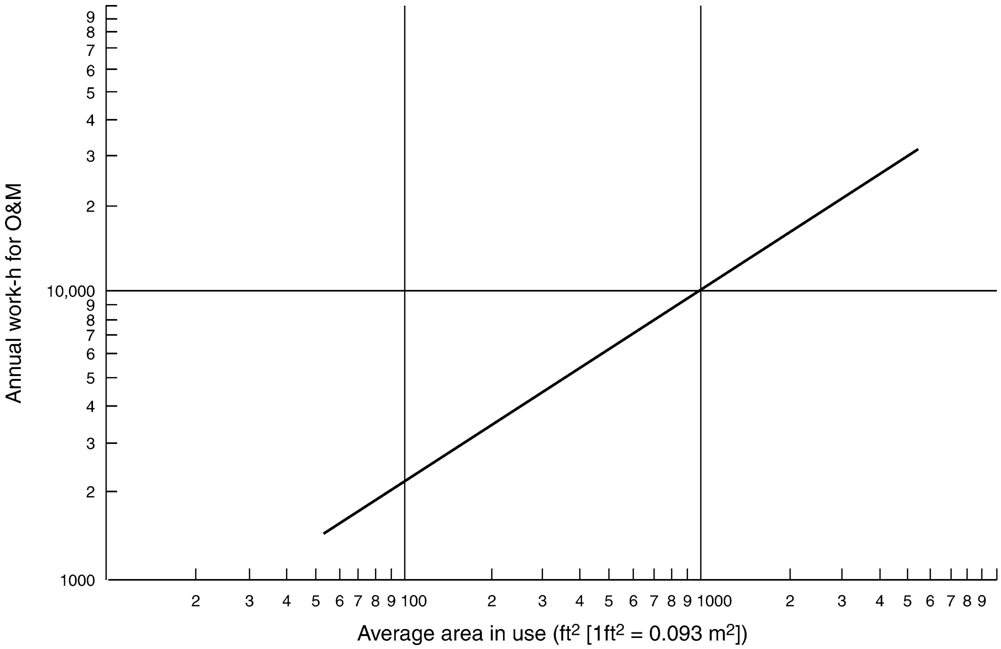


Fig. 10. Annual OM man-hour requirements-rotary drum vacuum filters (*Source:* US EPA).

this number is based on a June 1975 cost, it must be adjusted to the current 2006 design period. Using the Cost Index for Utilities (Appendix A), the cost would be multiplied by a factor of 528.12/190.49 = 2.77 (32). Hence, the yearly materials cost for a filtration area of 400 ft^2 (37.2 m^2) would be 11,080 USD in 2006 USD.

10. EXAMPLE

A 100,000 gpd biosolids stream at 5% solids (after addition of conditioning chemicals) will be fed at 4 $lb/ft^2/h$ for 16 h/d. Solids capture is 96% and cake solids are 25%. Determine:

a. Required filtration area.
b. Daily energy requirement.
c. Energy requirement per ton of captured solids.
d. Amount of water remaining in cake.

Solution

a. Daily weight of fed solids =100,000 gal/d × 8.34 lb/d × 0.05 = 41,700 lb/d.

Required filtration area = (41,700 lb/d)/(16 h/d × 4 $lb/ft^2/h$) = 651 ft^2.

b. From Fig. 11, for a surface area of 651 ft^2, the energy requirement = 490,000 kWh/yr
Daily energy requirement = 490,000 kWh/yr × (16/24)/365 d/yr.

= 895 kWh/d.

c. Energy requirement per ton of captured solids = (895 kWh/d × 2000 lb/t)/(41,700 lb/d × 0.96).

= 44.7 kWh/t.

d. Water remaining in cake = 41,700 × 0.96 × [(10.25)/0.25].

= 120,100 lb.

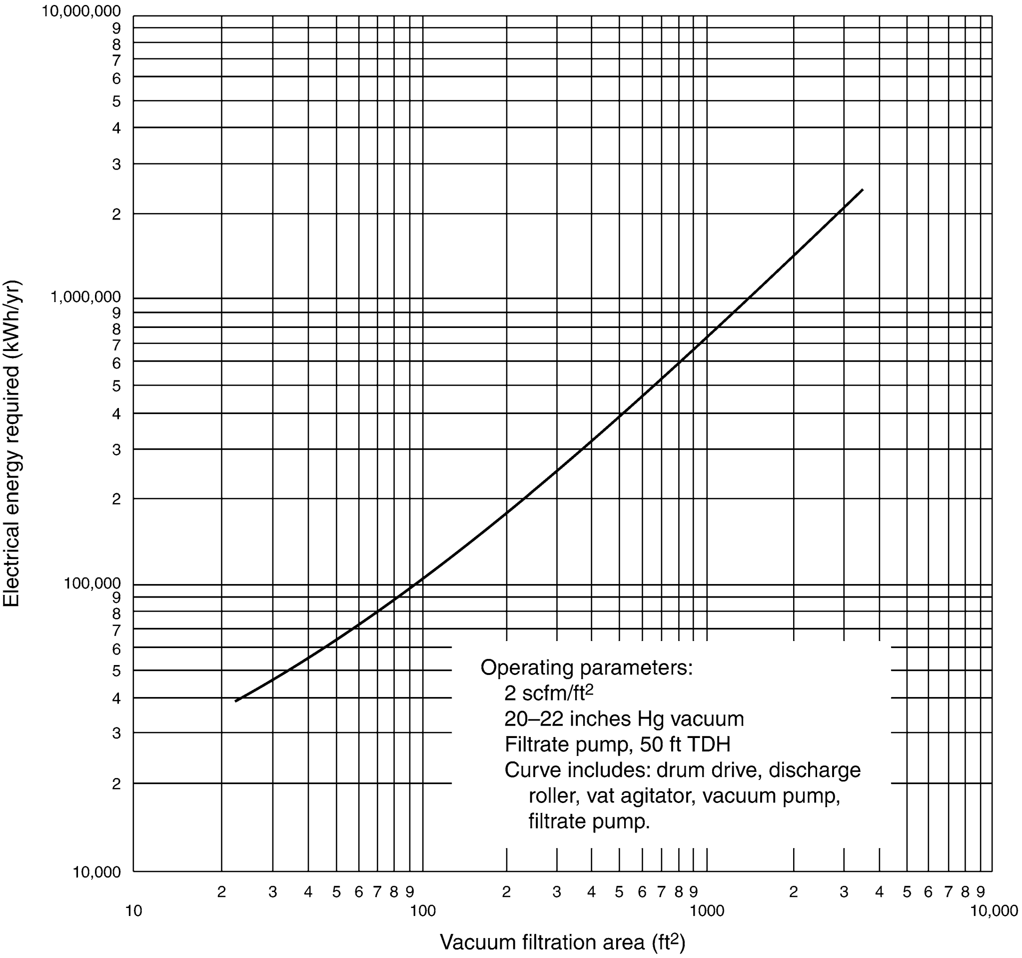


Fig. 11. Power consumed by rotary drum vacuum filtration process (1 ft^2 = 0.093 m^3) (*Source:* US EPA).

Table 7
Energy Requirements of Treatment Operations as Percent of Total Plant Energy

Treatment operation	% of total plant energy
Raw water pumping station	4.5
Headworks	0.4
Primary clarifiers	10.3
Activated sludge	55.6
Secondary clarifiers	3.7
Thickeners	1.6
Effluent filters	0.9
Utility water	3.6
Biosolids dewatering	7
Heating	7.1
Lighting	2.2
Post aeration/Chlorine mixer	3.1

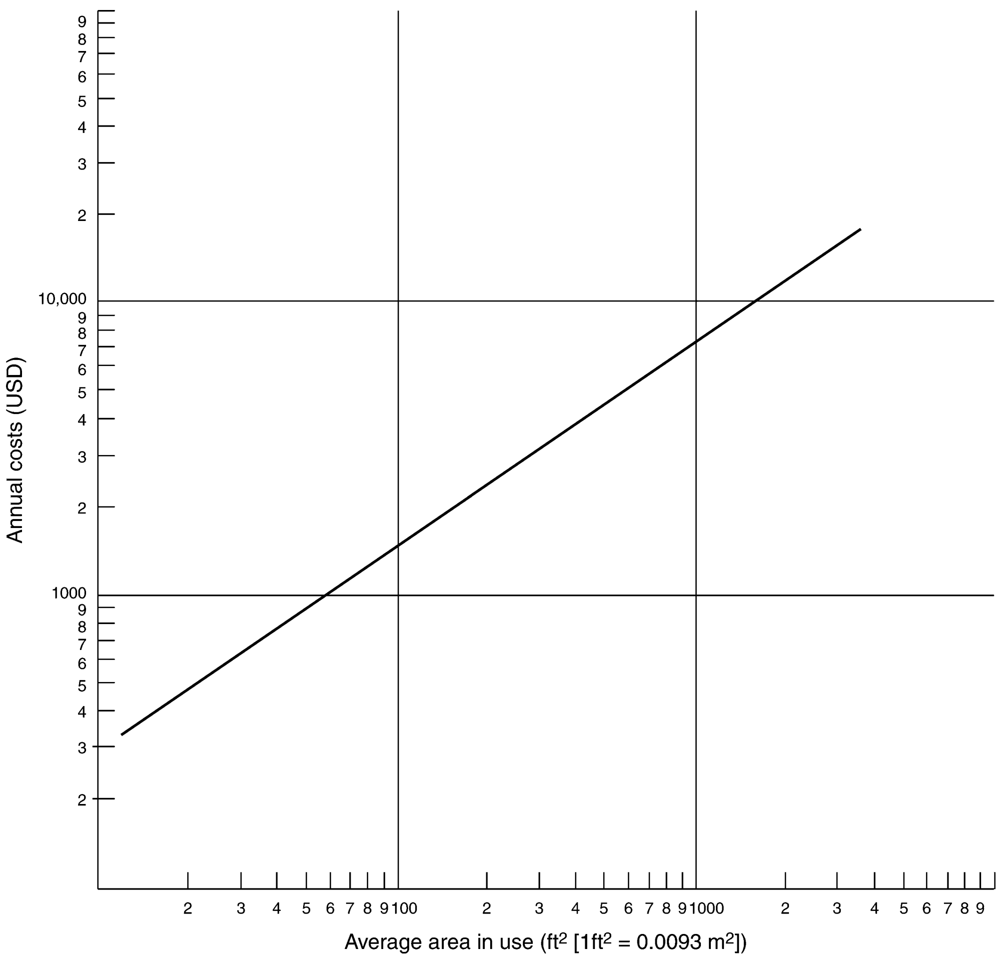


Fig. 12. 1975 Annual maintenance material cost for rotary drum vacuum filters (*Source:* US EPA).

NOMENCLATURE

A	Bed area
K	Bed permeability
L	Medium thickness
Q	Flow rate
R	Medium resistance
t	Ton (english unit)
T	Tonne (metric unit)
Δp	Driving pressure
μ	Filtrate viscosity

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APPENDIX

United States Yearly Average Cost Index for Utilities, US Army Corps of Engineers[a]

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

[a] From ref. 32.

17
Belt Filter Presses

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

Belt filter presses are used to remove water from liquid wastewater residuals and produce a nonliquid material referred to as "cake." Dewatered residuals, or cake, vary in consistency from that of custard to moist soil. Dewatering serves the following purposes (1):

a. Reducing the volume, thus reducing storage and transportation costs.
b. Eliminating free liquids before landfill disposal.
c. Reducing fuel requirements if residuals are to be incinerated or dried.
d. Producing a material which will have sufficient void space and volatile solids for composting when blended with a bulking agent.
e. Avoiding the potential of biosolids pooling and runoff associated with liquid land application.
f. Optimizing subsequent processes such as thermal drying.

Belt filter presses use single or double moving belts to dewater biosolids and other sludges continuously. The early belt presses used in the United States were those developed by Klein and by Smith and Loveless in the 1960s (2,3). Belt filter presses are currently very popular not only in the United States (4,5) but in other parts of the world as well (6).

From: *Handbook of Environmental Engineering, Volume 6: Biosolids Treatment Processes*
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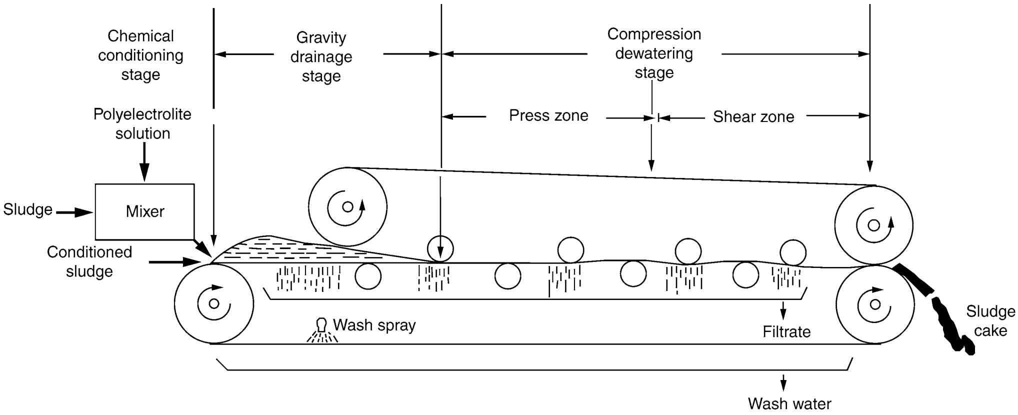


Fig. 1. The three basic stages of a belt press (*Source:* US EPA [9]).

2. PRINCIPLES OF BELT FILTERS

A belt filter dewaters by applying pressure to the biosolids to squeeze out the water. Biosolids sandwiched between two-tensioned porous belts are passed over and under rollers of various diameters. Increased pressure is created as the belt passes over rollers which decrease in diameter (7). Many designs of belt filtration processes are available, but all incorporate the following basic features: polymer conditioning stage, gravity drainage stage, and compression dewatering stage-low pressure squeezing and high pressure squeezing zones (8,1). Some belt filters include the added feature of vacuum boxes in the free drainage zone. About 6 in. Hg vacuum are applied to obtain higher cake solids. A "second generation" of belt filters has extended shearing or pressure stages that produce substantial increases in cake solids, but are more costly. Advanced designs provide a large filtration area, additional rollers, and variable belt speeds that can increase cake solids by 5%.

The general mechanical components of a belt filter press include dewatering belts, rollers and bearings, belt tracking and tensioning system, controls and drives, and a belt washing system. Figure 1 depicts a simple belt press and shows the location of the three stages. Although present-day belt presses are more complex, they follow the same principles indicated in Fig. 1.

Good chemical conditioning is the key to successful and consistent performance of the belt filter press, as it is for other dewatering processes. This is fully discussed in Chapters 11 and 12. After conditioning, the readily drainable water is separated from the slurry by discharge of the conditioned material onto the moving belt in the gravity drainage section. Typically, 1 or 2 min are required for drainage. Following drainage, the biosolids will have been reduced in volume by about 50% and will have a solids concentration of 6–10%. "The formulation of an even surface cake at this point is essential to the successful operation of subsequent stages of the dewatering cycle. The even surface prevents uneven belt tension and distortion while the relative rigidity of the mass of biosolids allows further manipulation and gives maximum speed through the machine" (6).

The third stage of the belt press begins as soon as the biosolids are subjected to an increase in pressure, because of either the compression of the biosolids between the carrying belt and cover belt or the application of a vacuum on the carrying belt.

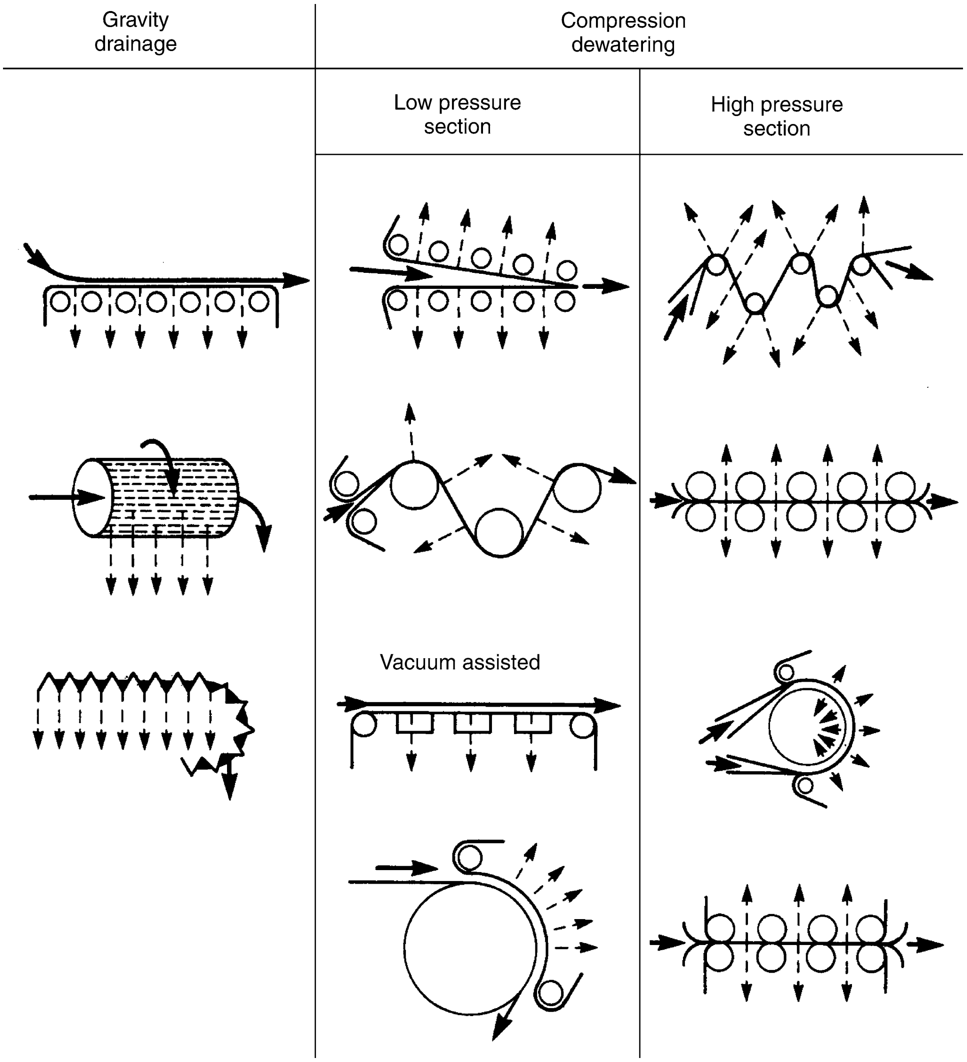


Fig. 2. Alternative design schemes of belt filter presses (*Source:* US EPA [9]).

Pressure can be widely varied by design, as shown by the alternatives on Fig. 2. During pressure application, the biosolids cake, squeezed between the two belts, is subjected to flexing in opposite directions as it passes over the various rollers. This action causes increased water release and allows more compaction of the sludge.

3. ADVANTAGES AND DISADVANTAGES

Advantages and disadvantages of belt filter presses for dewatering wastewater solids are summarized as follows:

1. Advantages
    a. High pressure machines are capable of producing very dry cake (9).
    b. Low power requirements (9).

   c. Staffing requirements are low, especially if the equipment is large enough to process the solids in one shift (10).
   d. Maintenance is relatively simple and can usually be completed by a wastewater treatment plant maintenance crew. Replacing the belt is the major maintenance cost (1).
   e. Belt presses can be started and shut down quickly compared with centrifuges, which require up to 1 h to build up speed (11).
   f. There is less noise associated with belt presses compared with centrifuges (11).
2. Disadvantages
   a. Odor may be a problem, but can be controlled with good ventilation systems and chemicals, such as potassium permanganate, to neutralize odor-causing compounds (12). Some manufacturers offer fully enclosed equipment to minimize odor and reduce vapor in the operating room air (13).
   b. Belt presses require more operator attention if the feed solids vary in their solids concentration or organic matter. This should not be a problem if the belt presses are fed from well-mixed digesters (11).
   c. Wastewater solids with higher concentrations of oil and grease can result in blinding the belt filter and lower solids content cake (1).
   d. Wastewater solids must be screened and/or ground to minimize the risk of sharp objects damaging the belt (1).
   e. Belt washing at the end of each shift, or more frequently, can be time consuming and require large amount of water (11). An automatic belt washing system and the use of effluent can minimize these costs.
   f. Machines are hydraulically limited in throughput (9).

4. APPLICATION AND PERFORMANCE

Belt filter presses can be used to dewater most biosolids generated at municipal wastewater treatment plants and are a common type of mechanical dewatering equipment (1,5,9). Using mechanical equipment to dewater solids might not be the most cost effective alternative for wastewater treatment plants operating at less than about 4 MGD. The selection of dewatering equipment should be based on the results of a site specific biosolids management plan which identifies both processing and end use of alternatives and estimates costs. It might be less expensive to haul liquid to an application site or pay a processing facility to dewater and process or landfill the dewatered cake. Smaller facilities should also evaluate nonmechanical dewatering methods, such as drying beds.

Manufacturers (14) should be consulted for design and performance data early in the planning stage. Data should be confirmed with other operating installations and/or pilot testing (1,15–17). Evaluation of equipment should consider capital and operating costs, including polymer, electricity, wash water, solids capture, and ventilation and odor control during dewatering and further processing or recycling. The operator can ensure system integration by requiring that the self-enclosed belt press, ventilation, and polymer system is supplied by a single provider. Because solids characteristics and quantity vary from plant to plant, it is important to evaluate different weaves, permeability, and solids retention abilities of dewatering belts to ensure optimum performance. Surveys of similar plants (18) or testing of wastewater solids can be helpful in the decision-making process.

It is difficult to generalize about the operating performance of belt presses because results depend on many factors: method of conditioning, maximum pressure, number

Table 1
Typical Dewatering Performance of Belt Filter Presses

Type of sludge	Feed solids (%)	Cake solids (%)	Polymer (pounds dry/t dry solids)
Raw primary (P)	3–10	28–44	2–9
WAS[a]	1–3	16–32	2–4
	0.5–1.5	12–28	4–12
P + WAS	3–6	20–35	2–10
P + trickling filter (TF)	3–6	20–40	3–10
Anaerobically digested			
P	4–10	26–36	2–6
WAS	3–4	18–22	4–8
P + WAS	3–9	18–44	3–9
Aerobically digested			
P + WAS	1–3	12–18	4–8
	6–8	20–30	2–5
Thermal conditioned			
P + WAS	4–8	38–50	0

Source: US EPA (9).
[a] WAS = Waste-activated sludge.
1 lb/t = 0.5 kg/T.

of rollers, and so on. Table 1 displays the range of performance of a high pressure belt press on various types of wastewater solids.

Medford, New Jersey (19) reported on a belt press dewatering aerobically digested biosolids from a contact stabilization system. Feed sludge of a 3–4% solids concentration was dewatered to a cake of 17–19% solids (19). Polymer was added for conditioning at 7–10 lb of dry polymer/t of dry feed solids (3.5–5 kg/T). The solids concentration in the combination wash water and filtrate was 100 mg/L for an overall solids capture of 99%. Komline-Sanderson (20) reported belt filter presses that received slurries ranging from 1 to 4% feed solids produced final products of 12–35% cake solids.

5. DESIGN CRITERIA

Belt presses are sized on the basis of weight or volume of solids to be dewatered rather than the wastewater flow to the plant. To determine how many presses are needed, the wastewater treatment plant must (1):

a. Determine the amount of primary solids that will flow through the plant per day.
b. Determine the amount of waste-activated or trickling filter solids produced per day.
c. Determine the volume of thickened solids to be dewatered per day.
d. Estimate the range of dry solids concentration in the feed.
e. Estimate future increases in solids production.
f. Anticipate changes in sewer discharges or operation that could change solids quality or organic matter content.

An effective biosolids management plan will include the afore-mentioned information. It is important to design for excess capacity so that the anticipated amount of incoming solids can be easily dewatered during operating hours. Allowing for excess capacity also ensures that the plant will not experience a build-up of solids if a unit is out of service. If only one unit is out of service, the plant should have an alternate program to remove solids in liquid form.

The polymer conditioning zone can be a small tank, approx 265–379 L (70–100 gal) located 0.6–1.8 m (2–6 ft) from the press, a rotating drum attached to the top of the press, or an in-line injector. The press manufacturer usually supplies this zone along with the belt filter press (10). The gravity drainage zone is a flat or slightly inclined belt unique to each model. Solids are dewatered by the gravity drainage of the free water. A 5–10% increase in solids concentration from the original biosolids should occur in this zone (10). The free water drainage is a function of wastewater solids type, quality, conditioning, screen mesh, and design of the drainage zone.

The low-pressure zone is the area where the upper and lower belts come together with the wastewater solids in between. This is sometimes called the "wedge zone," because the feed solids are sandwiched between the upper and lower belts. The low-pressure zone prepares the biosolids by forming a firm cake which can withstand the forces of the high pressure zone.

In the high-pressure zone, forces are exerted on the solids by the movement of the upper and lower belts as they move over and under a series of rollers of decreasing diameter. This produces a drier cake, an important factor for plants that incinerate the final product or face high end use or disposal costs. A biosolids management plan should evaluate the advantages and disadvantages of a high performance belt filter press.

An additional design feature is a self-enclosed facility to reduce odor and protect workers health (13). Workers in the belt press areas are exposed to aerosols from wash spray nozzles and pathogens and hazardous gasses such as hydrogen sulfide. Enclosing the press reduces visibility to the operators and produces a corrosive environment for the rollers and bearings, but automating the system can alleviate these problems.

The automation of belt presses is the subject of a Water Environment Research Foundation project (1). Benefits of automation include optimization of nonlinear variables which was rarely possible with manual or semiautomated operation, and the ability to produce dewatered cake at a constant rate. Generally, automation increases capital costs by 10%. Manufacturers claim that this extra expense is worthwhile because it lowers labor costs, reduces polymer use, and maximizes the solids content of the cake, reducing disposal and end use costs (21–23).

The choice of dewatering technique and chemical polymer or salts impacts dewaterability as well as the potential for odor during further processing or recycling to land.

Ancillary equipment for efficient operation of a belt press are as follows:

a. Polymer.
b. Mixing, aging, feed, liquid feed day tank.
c. Liquid residuals feed pump.
d. Odor control and ventilation.
e. Conveyor and/or pump to move dewatered cake.
f. An enclosed area to load trucks or containers.

Solids must be conditioned with polymer to ensure optimum performance. Polymer feed points should be designed at several locations to ensure flexibility and optimum performance. The solids/polymer mixture should be subject to gentle mixing as turbulent conditions can sheer the floc, minimizing polymer effectiveness. Polymer dilution and aging systems should be large enough to optimize polymer usage.

Table 2
Design Criteria for Belt Filter Presses

Sludge type	Conditioning	Solids to pressure filter (%)	Typical cycle length (h)	Solids filter cake (%)
Primary	5% FeCl$_3$, 10% Lime	5	2	45
Primary + FeCl$_3$,	10% Lime	4[a]	4	40
Primary + 2 stage high lime	None	7.5	1.5	50
Primary + WAS	5% FeCl$_3$, 10% Lime	8[a]	2.5	45
Primary + (WAS + FeCl$_3$)	5% FeCl$_3$, 10% Lime	8[a]	3	45
(Primary + FeC1$_3$) + WAS	10% Lime	3.5[a]	4	40
WAS	7.5% FeCl$_3$, 15% Lime	5[a]	2.5	45
WAS + FeCl$_3$	5% FeCl$_3$, 10% Lime	5[a]	3.5	45
Digested primary	5% FeCl$_3$, 10% Lime	8	2	45
Digested primary + WAS	7.5% FeCl$_3$, 15% Line	6–8[a]	2.5	45
Digested primary + (WAS + FeCl$_3$)	5% FeCl$_3$, 10% Lime	6–8[a]	3	40
Tertiary alum	10% Lime	4[a]	6	35
Tertiary low lime	None	8[a]	1.5	55

Source: US EPA (24).

[a] Thickening used to achieve this solids concentration.

Potassium permanganate or other oxidizing agents are often added to solids before dewatering. These have been shown to reduce odor caused by sulfides, reduce the amount of polymer needed, and increase cake solids content (12).

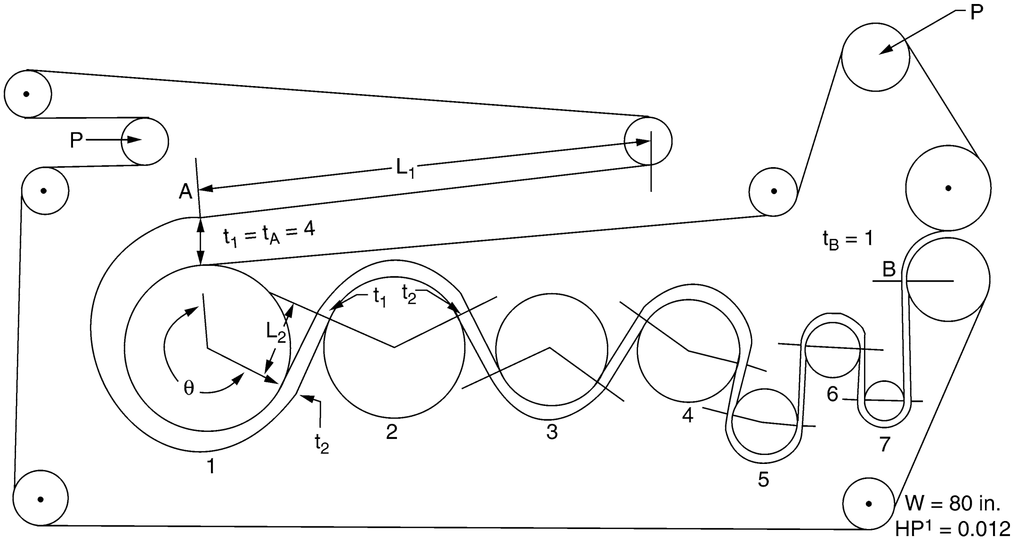
The following biosolids loadings are based on active belt area (9):

1. Raw primary
   a. Sludge loading = 27–34 gal/ft^2/h.
   b. Dry solids loading = 13.5–17 lb/ft^2/h.
2. Digested primary
   a. Sludge loading = 20–24 gal/ft^2/h.
   b. Dry solids loading = 20.5–24 lb/ft^2/h.
3. Digested mixed primary/secondary
   a. Sludge loading = 13–17 gal/ft^2/h.
   b. Dry solids loading = 6.7–8.4 lb/ft^2/h.

Typical conditioning requirements, cycle times, and cake solids for various sludge types are presented in Table 2.

6. DESIGN OF HIGH PRESSURE ZONE

The high pressure zone is critical to good belt press performance (high cake solids and recoveries). A procedure summarized by EPA (10) contains models to describe the various effects in a typical high pressure zone (*see* Fig. 3). US EPA provided these models to familiarize the engineer with the belt filter press design process. The models can be used to calculate the following parameters:



Roll	1	2	3	4	5	6	7
Diameter, in.	36	30	24	21	13	10	8
L_1, in.	94	16	15	16	11	17	10
L_2, in.	16	15	16	11	17	10	24
θ	244°	130°	114°	160°	185°	180°	180°
$\frac{\pi D\theta}{360}$	76.6	34	23.9	29.3	21	15.7	12.6

Fig. 3. Typical high pressure zone (*Source:* US EPA [10]).

1. ***Pressure on the biosolids cake because of drive torque*** (force required to pull the belt through the press):

$$\text{psi}_1 = \frac{2F_1}{D} = \frac{1700\,\text{HP}^1}{D}(\text{fpm}) \tag{1}$$

where psi_1 is the maximum pressure on the biosolids cake because of F_1; F_1 is the lb of force resulting from drive torque per inch of belt width; D is the roller diameter (in.); HP^1 is the drive horsepower per inch of belt width per belt; fpm is the belt speed (ft/min).

2. ***Pressure on the sludge cake because of belt tensioning*** (for presses that use pneumatic or hydraulic cylinders to tension the belts):

$$\text{psi}_2 = \frac{2F_2}{D} + 2P\cos\frac{a}{DWY/2} \tag{2}$$

where psi_2 is the average pressure on sludge cake because of F_2; F_2 is the lb of force resulting from take-up tension per inch of belt width-required to prevent slack belts and to provide traction for the drive rolls; P is the resultant force from tensioning roller actuator. It is the pressure (force) you set and can easily measure; a is the angle between belt force resultant and actuating cylinder axis; D is the diameter of roller (in.); W is the active belt width (in.); Y is the belt wrap angle at take-up roller.

3. ***Pressure on the sludge cake owing to belt elasticity:***

$$\text{psi}_3 = \frac{2F_3}{D} \tag{3}$$

where psi_3 is the average pressure on the sludge cake because of F_3; F_3 is the lb of force owing to belt elasticity per inch of belt width (where $F_3 = 2eE/D$); D is the roller diameter (in.); E is the modulus of elasticity of the belt (i.e., stress/strain before yield point); e is the belt strain (Δ/L_1); Δ is the belt stretch ($L_0–L_i$) (cm or in.); L_1 is the tangent length of belt entering roller; L_0 is the length of outer belt around roller between tangent points on adjacent rollers; L_i is the length of inner belt around roller between tangent points on adjacent rollers.

4. ***Total pressure on the cake at any roller:***

$$\mathrm{psi} = \mathrm{psi}_1 + \mathrm{psi}_2 + \mathrm{psi}_3 \tag{4}$$

or

$$\mathrm{psi} = \frac{2(F_1 + F_2 + F_3)}{D}$$

With these equations the engineer can calculate the total pressure on the biosolids cake at each roller to ensure that there is a gradually increasing pressure on each successive roller. These equations also allow determination of roller and shaft diameters and bearing size requirements, which can then be compared with the belt press manufacturer's specifications.

The parameters needed to evaluate a design are as follows:

a. D = Diameter of each roller (in.).
b. L_1 = Tangent length of belt entering each roller.
c. L_2 = Tangent length of belt leaving each roller.
d. θ = Angle of wrap of the belt around the roller.

The afore-mentioned parameters are available from the manufacturer's specifications. Other parameters are as follows:

a. Q = Biosolids throughput rate (lb of solids/min).
b. t_A = Cake thickness at the entrance to the high pressure zone (in.).
c. t_B = Cake thickness at exit.
d. C_A = Cake solids concentration at the entrance to the high pressure zone as a decimal.
e. C_B = Cake solids concentration at the exit of the high pressure zone as a decimal.
f. E = Modulus of elasticity from belt manufacturer's specifications (lb/in^2).
g. HP^1= Drive horsepower per inch of belt width per belt.
h. W = Belt width (in.).
i. t_1 = Cake thickness at entrance of roller (in.).
j. t_2 = Cake thickness at exit of roller (in.).
k. t_a = Average cake thickness at roller, $(t_1 + t_2)/2$.
l. L = Effective length of belt (portion over each drum) ($D\theta/360$).

An example using these equations to evaluate a belt filter press design is shown in Section 10.

7. ODOR CONTROL

Odor complaints at wastewater treatment plants and biosolids end use sites can interfere with implementation of the most cost effective biosolids management options. Odor control measures should be included when designing dewatering facilities. Odor control is addressed in more detail in another fact sheet (25), but briefly, the methods include:

a. Using a self enclosed belt press.
b. Adding potassium permanganate or other oxidizing agent to minimize odor in the solids.
c. Minimizing liquid storage before belt pressing to less than 24 h. The longer the solids are stored, the lower the pH, the higher the liquid ammonia concentration, and the higher the organic sulfide emissions (26).
d. Conducting bench-scale and full-scale testing of liquid sludge to determine if combined storage of primary and waste-activated sludges (WAS) accelerates the deterioration of biosolids (26).
e. Specifying polymers that are stable at elevated temperatures and pH. This is especially important at facilities using lime stabilization or high temperature processing such as heat drying, thermophyllic digestion, or composting.

The main purpose of a self-enclosed system is to minimize the amount of foul air needing treatment in an odor control system. An induced draft fan provides a slight negative pressure. The system design should:

a. Minimize gaps in the enclosure.
b. Minimize enclosure volume.
c. Locate mechanical and electrical components requiring maintenance outside the enclosed area for easy access and reduced maintenance.
d. Include automation to optimize performance of the belt press.
e. Use stainless steel materials.
f. Provide multiple access hatches to allow operator viewing and clean up.
g. Provide for easy removal of the belt for replacement (13).

8. OPERATION AND MAINTENANCE

It is important to monitor operating parameters to achieve optimum performance and ensure that solids are properly conditioned and that good gravity drainage occurs. The manufacturer should provide operation and maintenance training after installation as well as ongoing training to maintain skills.

A belt press operator is responsible for polymer mixing, dosing and monitoring usage, and observing the feed and cake several times per day, making adjustments as necessary. Failure of the chemical conditioning process to adjust to changing sludge or biosolids characteristics can cause operational problems (27). If it is underconditioned, sludge does not drain well in the gravity drainage section, and the result is either extrusion of inadequately drained solids from the compression section, or uncontrolled overflow of sludge from the drainage section. Both underconditioned and overconditioned sludges can blind the filter media. In addition, overconditioned sludge drains so rapidly that solids cannot distribute across the media. Inclusion of a sludge blending tank step before the belt press reduces this problem.

Dewatering belts should be designed for easy replacement with minimum downtime. Belt washing should occur daily after the cake is removed. Replacement of filter belts is a common maintenance requirement. Belt life averages about 2700 running hours, with a range of 400–12,000 h (1).

The combined filtrate and belt wash water flow is normally about 1.5 times the incoming flow. Some belt presses recirculate wash water from the filtrate collection system, but normally, secondary effluent or potable water is used. This total flow contains between 100 and 1000 mg/L of suspended solids and is typically returned either to the primary or secondary treatment system.

Belt presses have numerous moving parts, and spare parts should be kept available to prevent prolonged unit down-time. Belts, bearings, and rollers require frequent lubrication and could deteriorate quickly, especially in municipal wastewater treatment plants where normally preventive maintenance is not practised.

It is important for the operator to keep records of all press performance parameters, including the volume of biosolids fed to the press, polymer dosage, and potassium permanganate or other chemical usage. A sample of the biosolids to the press, cake discharge, and filtrate should be taken at least once per shift and analyzed for total solids. At the end of each shift, the belt should be cleaned with high-pressure wash water. Labor is relative to plant size. A plant with a single belt press needs 4–8 staff h/d (including lab testing), whereas 6–8 presses can be operated with 8–10 staff h/d. Large plants use less operating effort per belt press. Highly automated systems reduce labor requirements, but require an instrumentation specialist to maintain the system.

Control of a belt filter press system includes the following components and operations (28):

1. **Belt tracking.** Correct tracking of the filter belt is very important to assure minimum wear and damage to the belts. Some units are equipped with automatic adjusting devices designed to correct roller alignment automatically. Other units require a periodic check and adjustment, if necessary, to be made by the plant operator.
2. **Spray adjustment.** Correct adjustment of spray nozzles used to clean the underside of the belt or screen is also important. Biosolids buildup on the underside of the belt creates a tracking problem. Just enough spray should be used so that the underside of the belt remains clean.
3. **Inspection.** The filtrate should be relatively clear and no excessive biosolids buildup should be occurring anywhere along the belt or rollers. Once the operator is familiar with this equipment it should be possible to judge the operation of the belt filtration unit by visual appearance.
4. **Sampling and analysis.** Influent biosolids and filtrate samples might be obtained through the valves provided. Dewatered biosolids samples may be obtained after the biosolids have been removed from the belt by the scraper mechanism. Samples should be analyzed according to procedures specified in *Standard Methods.*
5. **Biosolids conditioning.** Proper biosolids conditioning is an important step in any dewatering process. Biosolids conditioning results in flocculation of the small solid particles into larger particles which have enough size and strength to bridge the openings in the filter belt and, thus, be retained on that belt. In order to determine the best chemicals and chemical dosages to use, jar testing should be performed on several biosolids samples. The optimum dosage will be, above which little or no increase in floc size or supernatant clarity is noted. In addition to the chemicals, the following parameters will affect the final percent solids concentration obtained by the belt filtration unit:
    a. Incoming biosolids percent concentration.
    b. Loading or application rate (lb/h) of biosolids to belt filtration unit.
    c. Operating speed of belt filtration unit.
    d. Compression of the pressure rollers.
6. **Solids.** In general, thicker incoming biosolids will produce a drier cake. However, varying the initial solids concentration is not normally used as a process control variable. It is customary, unless special conditions apply, to deliver as thick biosolids as practical to the belt filtration unit.

7. **Loading rate.** The biosolids loading rate or application rate has a significant effect on the performance of the belt filtration unit. A loading rate that is too high will cause poor performance. The ideal loading rate is the highest rate at which the system can be run without a drop in the desired performance. This rate is dependent on the rate of travel of the filter belt.
8. **Belt rate of travel.** The speed of the filter belt should be increased along with a corresponding increase in the rate of biosolids feed. The exact speed at which the unit should be operated depends on the results desired in terms of biosolids cake dryness, of percent biosolids retained on the filter belt, and the dewatering rate of the biosolids. This speed can only be determined by trial and error operation. Once this setting has been determined, infrequent minor adjustments should be required. If the unit is to be shut down, these settings should be noted for use when restarting.
9. **Compression.** Determination of the best compression of the pressure rollers may require a certain amount of experimentation through actual operation to set properly. Once set they should require little adjustment.

9. COSTS

Capital costs for belt filter presses vary with the size of the equipment. According to a recent US EPA Fact Sheet (1), vendor estimates vary from 53,500 USD (2006 USD) (0.5 m belt, approximate capacity of 500 dry lb/h) to 129,000 USD (1.5 m belt with approximate capacity of 1625 dry lb/h). These cost estimates are based on a feed material which is 5% solids and were updated from year 2000 to reflect year 2006 costs using the Cost Index for Utilities (Appendix); costs were multiplied by a factor of 528.12/468.05 = 1.13 (29). These prices do not include the cost of installation, shipping, or ancillary equipment, such as flow control and centrate management.

Cost information is also available from another US EPA publication (30). According to this publication, construction costs for a belt filter press, sludge feed pump, polymer pump, and control panel to dewater 1000 lb (454 kg) of sludge/h was 237,000 USD (2006 USD). To dewater 2500 lb (1134 kg)/h, the cost would be 293,000 USD. These costs were updated from 1977 to reflect the 2006 costs using the Cost Index for Utilities (Appendix).

The electrical power requirement in kWh/yr to continuously operate a filter press of various volumes is given in Fig. 4.

Table 3 lists labor requirements for the operation and maintenance of belt filter presses. The labor indicated includes periodic operational adjustments and minor routine maintenance.

Overall operation and maintenance costs range from 90 USD (2006 USD)/dry ton of solids to 225 USD/dry t. Typical polymer conditioning costs for belt filter press dewatering range from 2.86 USD/MG to 98.4 USD/MG, and average 26.3 USD/MG. Permanganate adds about 1.08 USD/MG to the cost of dewatering the biosolids. These costs vary widely, depending on the source of the residuals. The polymer costs for raw primary may cost 13.6 USD/dry t of solids but may be as high as 90 USD/dry t for residuals that are difficult to dewater (1).

Costs of construction, operation and maintenance for various belt press capacities are shown in Table 4.

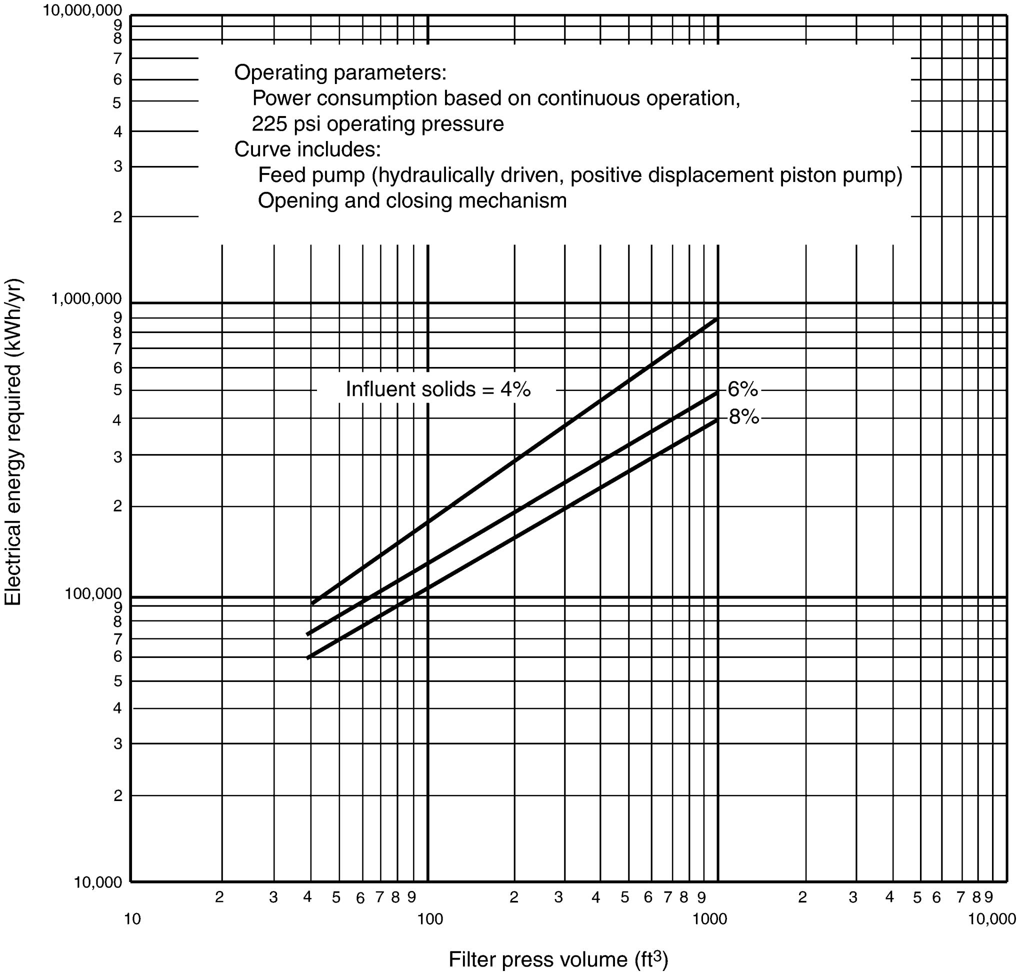


Fig. 4. Electrical energy requirements of belt filter presses (*Source:* US EPA [24]).

Table 3
Labor Requirements for Belt Filter Presses

	Labor (h/yr)		
Number of units	Operation	Maintenance	Total
1	265	100	365
2	530	200	730
3	795	300	1095
4	1060	400	1460
5	1325	500	1825

Source: US EPA (28).

Table 4
Costs for Construction, Operation, and Maintenance of Belt Filter Presses

Plant sludge loading (lb dry solids/d)	Machine capacity (lb dry solids/d)	Construction cost (2006 USD)	Power cost (2006 USD)	Labor cost (2006 USD)	Maintenance cost (2006 USD)
16,000	24,000	237,000	1980	27,000	3440
40,000	60,000	293,000	2920	27,000	4170
66,000	99,000	403,000	4170	27,000	5630

Source: US EPA (30).

The assumptions used for calculating the costs shown in Table 4 are as follows:

a. Costs were updated from 1977 to reflect the 2006 costs using the Cost Index for Utilities (Appendix A); costs were multiplied by a factor of 528.12/215.84 = 2.45 (29).
b. Type of sludge: Primary and secondary (anaerobically digested)—5% solids concentration.
c. Sludge production: 7 d/wk.
d. Dewatering operation: 7 d/wk; 16 h/d.
e. Construction costs include belt filter press, sludge feed pumps, polymer pumps, and control panels.
f. Labor cost: 35 USD/man-h.
g. Power cost: 0.047 USD/kW.

10. DESIGN EXAMPLES

10.1. Design Example 1

The designer for an existing wastewater treatment plant has calculated that the plant needs to dewater 5000 lb of dry sludge (2268 kg)/d, 5 d/wk. The sludge to be dewatered is a mixture of one part primary and two parts waste-activated, stabilized by a two-stage, high-rate, anaerobic digestion process. Total feed solids concentration to the belt filter press was 2.8%. Determine the solids capture in percent knowing that pilot plant testing with a 1-m-wide belt filter press produced the following results as follows:

a. Total solids in the dewatered sludge ranged from 23 to 30%, averaging 25%.
b. Optimum polymer dosage was 6–8 lb of dry polymer/t (3–4 kg/T) of dry feed solids, or 80–100 lb of liquid polymer/t (40–50 kg/T) of dry feed solids.
c. At the optimum polymer dosage, the total solids in the filtrate plus wash water flow were 2000 mg/L. The suspended solids averaged 900 mg/L.
d. Optimum hydraulic feed rate at 2.8% solids for 1-m-wide belt was 47 gpm (3 L/s).
e. Washwater requirements were 25 gpm (1.6 L/s).

Solution

On the basis of pilot plant data, the engineer decided that a single 1-m-wide belt filter press could dewater the 5000 lb (2268 kg) of sludge in 7.6 h. As it was important that the wastewater treatment plant always be able to dewater sludge, two 1-m-wide belt filter presses would be purchased.

The cost of dry polymer in 50 lb (22.7 kg) bags was 4.33USD/lb; for liquid polymer in 55 gal, 650 lb (208 L295 kg) drums, the cost was 0.30 USD/lb.

Daily cost for dry polymer at 8 lb/t (4 kg/T) would be:

5000 lb solids/d × 8 lb polymer/2000 lb solids × 4.30 USD/lb polymer = 86 USD/d.

Daily cost for liquid polymer at 100 lb/t (50 kg/t) would be:

5000 lb solids/d × 100 lb polymer/2000 lb solids × 0.30 USD/lb polymer = 75 USD/d.

Because sludge characteristics can change with time, a dual polymer system capable of utilizing either liquid or dry polymer will be installed. Because liquid polymer is currently less expensive, it will be used initially.

To allow subsequent computation of solids capture, the filtrate flow is calculated, using a suspended solids balance and a flow balance. The specific gravity of the feed, dewatered cake, and filtrate are assumed to be 1.02, 1.07, and 1.01, respectively.

The suspended solids balance is:

$$(47 \text{ gal feed/min}) (8.34 \times 1.02 \text{ lb feed/gal feed}) (0.028 \text{ lb solids/lb feed})$$
$$= (Q \text{ gal filtrate/min}) (8.34 \times 1.01 \text{ lb filtrate/gal filtrate}) (900 \text{ lb solids}/10^6 \text{ lb filtrate})$$
$$+ (M \text{ gal cake/min}) (8.34 \times 1.07 \text{ lb sludge/gal cake}) (0.25 \text{ lb solids/lb cake}).$$

The flow balance is:

$$47 \text{ gal feed/min} + 25 \text{ gal washwater/min} = Q \text{ gal filtrate/min} + M \text{ gal cake/min}.$$

The suspended solids and flow balances are solved simultaneously:

The flow of filtrate (Q) = 67.2 gpm (254 L/min)

The flow of cake (M) = 72 − 67.2 = 4.8 gpm (18 L/min)

Solids capture = [(Solids in feed—solids in filtrate)/solids in feed] × 100

$$= \left(\frac{47\,(8.34 \times 1.02) - 67.2\,(8.34 \times 1.01)\, 900/10^6}{47(8.34 \times 1.02)\,(0.028)} \right) \times 100$$

$$= 95\%.$$

All filtrate is returned to the secondary treatment process.

10.2. Design Example 2

A 100,000 gpd primary sludge stream at 5% solids (after addition of conditioning chemicals) will be fed to a filter press with a 2-h cycle time. Solids capture is 96% and cake solids are 45% (*see* Table 2). Operation is 16 h/d. Determine:

a. Pressing capacity.
b. Total energy requirement.
c. Energy requirement per ton of captured solids.
d. Weight of water remaining in cake.

Solution

1. Pressing capacity.
    a. Daily sludge flow = (100,000 gal/d)/7.48 gal/ft^3 = 13,369 ft^3/d.
    b. Number of cycles per day = (16 h/d)/2 h cycle = 8 cycles/d.
    c. Required pressing capacity = (13,369 ft^3/d)/8 cycles/d = 1670 ft^3.
2. Total Energy requirement.

Using two presses:
Capacity of each press = 1670 ft^3/2 = 835 ft^3.
From Fig. 4 (for 835 ft^3 press), energy requirement per press = 600,000 kWh/yr
Total energy requirement = 2 × 600,000 = 1,200,000 kWh/yr.

3. Energy requirement per ton of captured solids.
   lb of solids fed per day = 100,000 gal/d × 8.34 lb/gal × 0.05
   = 41,700 lb/d
   kWh/t of captured solids = (2192 kWh/d × 2000 lb/t)/41,700 × 0.96 = 109.5 kWh/t.
4. Weight of water remaining in cake.
   Water remaining in cake = (41,700 × 0.96) [(1 – 0.45)/0.45] = 48,928 lb.

10.3. Design Example 3

Evaluate the following belt filter press design. The magnitudes of the values for D, L_1, L_2, and θ are shown in Fig. 3. The magnitudes of the other parameters follow:

a. $W = 80$ in.
b. $Q = 125$ lb/min.
c. $t_A = 4$ in.
d. $C_A = 0.1$.
e. $C_b = 0.38$ (desired value).
f. $HP^1 = 0.0127$ HP/in.
g. fpm = $2.3\ Q/(W \times t_A \times C_A) = 9$ fpm.

Solution

a. Calculation of cake thickness at exit of high pressure zone.

$$t_B = (0.95 \times t_A \times C_A)/C_B = 1 \text{ in.} \tag{5}$$

This equation assumes 95% solids capture across the unit.

b. Plot: $F_1 + F_2$, in lb/in vs Dθ/360
c. Calculate:

$$e + \Delta/L_i = \frac{2t_a}{[360(L_1 + L_2)/n\theta] + D} \tag{6}$$

d. Calculate:

$$F_3 = 939e - \left(\frac{100}{2000e^2} + 1\right) + 108 \tag{7}$$

e. $\Sigma F = F_1 + F_2 + F_3$
g. $\text{psi} = \dfrac{2}{d\Sigma F}$

Comments and interpretation of results (*see* Table 5):

a. The value of t_A will have some maximum allowable value according to the press design.
b. t_a is the value at the center of the roller, halfway between tangent points.
c. The total pressure (psi) calculated for each roller is compared to the pressures specified by the manufacturer. In this example, the sharp increase in pressure between rollers 4 and 5 suggests that biosolids would either extrude into the belt mesh, clogging the belt or squeeze out from between the two belts. It is important to remember that the pressure

Table 5
Summary of Pressures Calculated for Each Roller

Roll	D	L_1	L_2	θ	$(F_1 + F_2)$	t_1	t_2	t_3	O	F_3	ΣF	psi
1	36	94	16	244	43	4	2.9	3.45	0.079	174	217	12
2	30	16	15	130	53	2.9	2.4	2.65	0.092	188	241	16
3	24	15	16	114	58	2.4	2.1	2.25	0.082	178	236	20
4	21	16	11	160	63	2.1	1.7	1.9	0.094	190	253	24
5	13	11	17	185	67	1.7	1.4	1.55	0.102	197	264	41
6	10	17	10	180	70	1.4	1.2	1.3	0.096	192	262	52
7	8	10	24	180	73	1.2	1	1.1	0.074	169	242	61

Source: US EPA (10).

should increase gradually from roller to roller and from zone to zone. Sharp increases of pressure can indicate operating problems.

d. In this example, the belt tension (ΣF) is more than 250 lb/in. in many cases. A force of this magnitude tends to deform most dewatering belts, which suggests problems with belt tracking and alignment as the belt wears.
e. The angle of wrap (θ) should be as close to 180° as possible.
f. Use F_1, F_2, and F_3 for analyzing the size of the rollers and bearings in the other zones. Failure to include F_3 when analyzing the high pressure section rollers and bearings could result in bearing and/or shaft failure.
g. The design engineer should require a submittal of the calculations from the manufacturer to confirm bearing and shaft design.
h. To calculate the reaction at each bearing for rollers in the high pressure zone with a bearing on each end of the roller, use Eq. (8):

$$R = W\,[2(F_1 + F_2) + F_3]\sin(\theta/2) \tag{8}$$

10.4. Design Example 4

A 0.22-m^3/s (5-MGD) activated sludge wastewater treatment plant with primary clarifiers is proposed for a town. The plant will be dewatering raw biosolids during an 8 h d, 5 d/wk. The biosolids will be thickened before dewatering to 5% solids. How many and what size belt presses will be required?

Solution

Belt presses are not sized on the basis of wastewater flow to the plant, but on the basis of the weight or volume of biosolids to be dewatered. The following calculations show how the required number of presses can be determined.

a. Determination of the amount of primary sludge.
   Influent total suspended solids concentration is determined by laboratory analysis of the wastewater that will be flowing through the plant. For this example, assume 220 mg/L,
   Total suspended solids = 220 mg/L × 5 MGD × 8.34 lb/MG/mg/L = 9180 lb/d (4168 kg/d).
   A primary clarifier will remove an average of 60 % of the suspended solids.
   Therefore: Total primary sludge = 0.6 × 9180 lb/d = 5508 lb/d (2501 kg/d).
   Assume the primary sludge is 3.5% solids,
   Total volume of primary sludge = (5508 lb/d)/[(0.035)(8.34 lb/gal)]
   = 18,858 gal/d (71.38 m^3/d).

b. Determination of the amount of WAS.
Influent BOD_5 is determined by analyzing the wastewater. For this example, assume 220 mg/L, a 30% removal of BOD_5 in the primary clarifier, and a total plant removal of 90% of the BOD_5,
BOD_5 removal = 0.7 (220 mg/L) – (0.1)(0.7)(220 mg/L) = 139 mg/L
BOD_5 removed = 139 mg/L × 5 MGD × 8.34 lb/MG/mg/L = 5780 lb/d (2622 kg/d)
Assume a biosolids yield coefficient of 0.5 at SRT of 16 d:
Solids production = 0.5 (5780 lb/d) = 2892 lb/d (1313 kg/d).
Total waste biosolids will then be equal to the amount of suspended solids in the primary effluent plus the solids production in the aeration system minus the effluent suspended solids,
Biosolids = 0.4 (9180 lb/d) + 2890 lb/d – 918 lb/d = 5646 lb/d (2563 kg/d).
Assume that WAS is 1% solids:
Total volume of biosolids = (5646 lb/d)/(0.01 × 8.34 lb/gal) = 67,657 gpd (256 m^3/d)

c. Determination of the volume of the 5% thickened biosolids to be dewatered per day.
Total solids (primary + WAS) = 5508 lb/d + 5646 lb/d = 11,154 lb/d.

$$\text{The total volume of the 5\% mixed biosolids} = \frac{11{,}154}{0.05 \times 8.34}$$

$$= 26{,}750 \text{ gpd}$$

The plant wants to dewater only 5 d/wk, therefore:
26,750 gpd × 7/5 = 37,450 gpd (143 m^3/d).

A typical belt filter press has a hydraulic loading of 40 gpm/m of belt width: 37,450 gal/40 gpm = 936 min = 15.6 h, say 16 h watering period.

Therefore, a 1-m belt press would dewater the biosolids produced at this plant in a 16-h d. However, biosolids are to be processed on an 8-h d and it is also important to design for excess capacity. Therefore, two 1.5-m units should be used. With both units operating, the biosolids would easily be dewatered in 8 h. Further, there is enough excess capacity so that if one unit were out of service, the other unit could process all of the biosolids in a 10.5-h d, thus preventing a build-up of biosolids in the plant.

NOMENCLATURE

a	Angle between belt force resultant and actuating cylinder axis (deg)
C_A	Cake solids concentration at the entrance to the high pressure zone as a decimal
C_B	Cake solids concentration at the exit of the high pressure zone as a decimal
D	Roller diameter (in.)
e	Belt strain (Δ/L_1)
E	Modulus of elasticity from belt manufacturer's specifications (lb/in.^{2})
fpm	Belt speed (ft/min)
F_1	lb of force because of drive torque per inch of belt width (lb/in.)
F_2	lb of force because of take-up tension per inch of belt width (lb/in.)
F_3	lb of force because of belt elasticity per inch of belt width, where F_3 = 2eE/D (lb/in.)
HP^1	Drive horsepower per inch of belt width per belt (hp/in./belt)
L	Effective length of belt (portion over each drum) ($D\theta/360$) (in.)
L_0	Length of outer belt around roller between tangent points on adjacent rollers (in.)
L_1	Tangent length of belt entering roller (in.)

L_2	Tangent length of belt leaving each roller (in.)
L_i	Length of inner belt around roller between tangent points on adjacent rollers (in.)
P	Resultant force from tensioning roller actuator. It is the pressure (force) you set and can easily measure
psi_1	Maximum pressure on the biosolids cake because of F_1 (lb/in.^{2})
psi_2	Average pressure on sludge cake because of F_2 (lb/in.^{2})
psi_3	Average pressure on the sludge cake because of F_3 (lb/in.^{2})
Q	Biosolids throughput rate (lb of solids/min)
T	tonne (Metric)
t	ton (English)
t_1	Cake thickness at entrance of roller (in.)
t_2	Cake thickness at exit of roller (in.)
t_a	Average cake thickness at roller ($[t_1 + t_2]/2$)
t_A	Cake thickness at the entrance to the high pressure zone (in.)
t_B	Cake thickness at exit (in.)
W	Belt width (in.)
Y	Belt wrap angle at take-up roller (deg)
Δ	Belt stretch ($L_0 - L_i$) (in.)
θ	Angle of wrap of the belt around the roller (deg)

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APPENDIX

US Yearly Average Cost Index for Utilities, US Army Corps of Engineers[a]

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

[a] From refs. 28,29.

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Pressure Filtration

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

Pressure filter presses are used to remove water from liquid wastewater residuals and produce a nonliquid material referred to as "cake." Dewatered cake varies in consistency from that of custard (12–15% solids) to moist soil (20–40% solids) (1). Filter presses for dewatering were first developed for industrial applications and, until the development of diaphragm presses, were only slightly modified for municipal applications.

The original or early models of the press were sometimes called plate and frame filters, because they consisted of alternative frames and plates on which filter media rests or is secured. The frames provide both structural integrity and spacing between the plates. The frames could be changed to provide different cake thicknesses. The unit had a fixed and a movable end, which promoted pressure maintenance during the filtration cycle (Fig. 1). There are few, if any, plate and frame units in service for municipal applications today, because this configuration is not particularly suitable for the filtration

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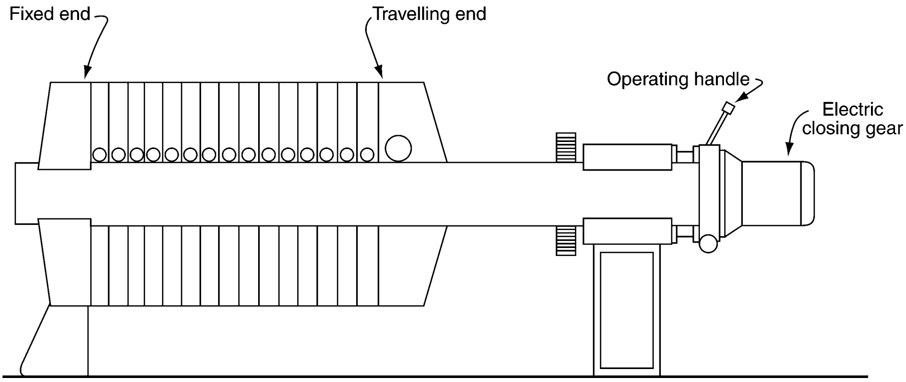

Fig. 1. Side view of a recessed plate pressure filter press (*Source:* US EPA).

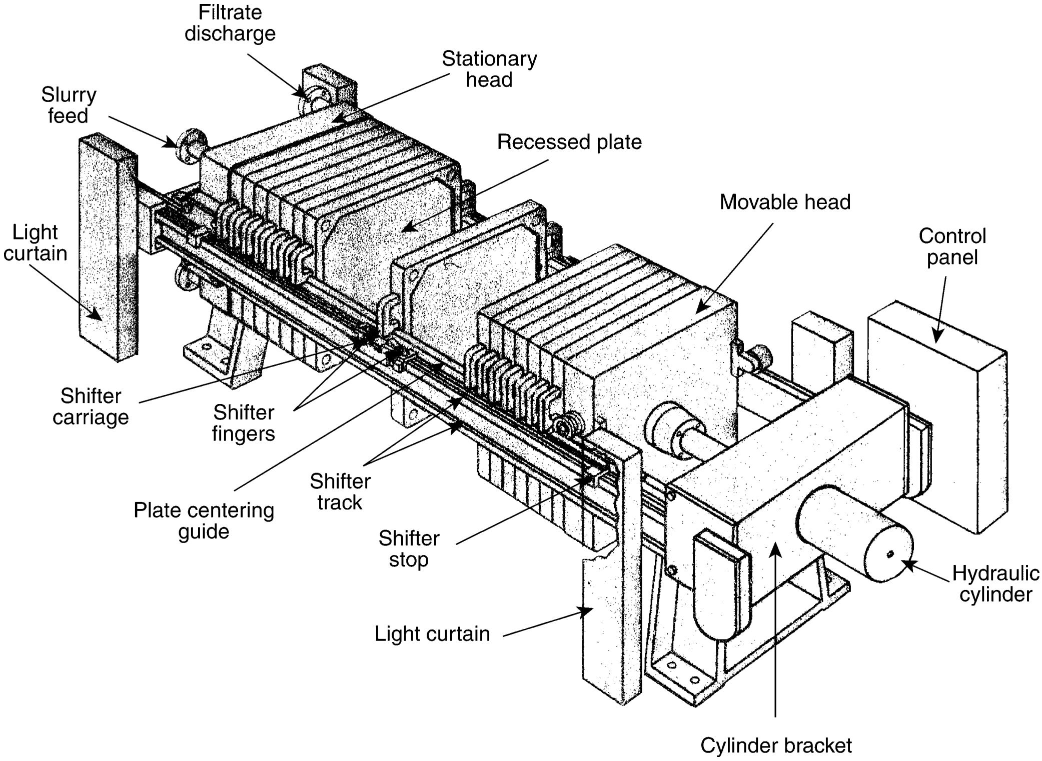

Fig. 2. Fixed-volume recessed plate pressure filter press (*Source:* US EPA).

of hydrous pseudoplastic materials like municipal biosolids. However, test filters sometimes are plate and frame, and they are used to determine the optimum cake thickness.

The equipment commonly in use for the dewatering of municipal biosolids falls into one of the two categories. The fixed-volume recessed plate filter and the variable-volume diaphragm filter press. A typical fixed-volume recessed plate filter press is shown in Fig. 2. Precoating the filter media or substantial chemical conditioning of biosolids is normally required. This is particularly true for such difficult-to-dewater materials as waste-activated sludge (WAS) or aerobically digested biosolids. A chemical conditioning station is therefore almost invariably a part of a facility that uses filter presses to dewater municipal biosolids.

There is usually a higher degree of operator activity associated with filter presses than with most other types of dewatering (2). As a result, filter presses, for the most part, have been employed in wastewater treatment facilities ≥25–50 MGD (1.1–2.2 m^3/s). On the other hand, this equipment can produce a very dry cake, probably the driest cake produced by conventional dewatering equipment (30–50% solids) (3,4). Hence, it has substantial attractiveness.

2. PROCESS DESCRIPTION

The recessed plate filter press shown in Fig. 2 consists of polypropylene squares, which may be 2–4 ft across, with a concave depression and a hole in the middle. Two plates are joined to create a chamber to pressurize solids and squeeze out liquid through a filter cloth lining the chamber. Several plates (ranging from 12 to 80, depending on the capacity required) are suspended from a frame face to face. A series of chambers is formed when the press is closed. Conditioned solids are pumped into the center hole to fill each chamber. As pressure increases, either by adding more conditioned solids (fixed volume press) or by expanding a membrane (variable volume press), solids are retained on the filter cloth while liquid passes through and is drained away from the machine. Free water is released and passes through the filter cloth during the filling phase. Pressure builds as the chamber fills with solids beginning the consolidation process. To perform effectively, the terminal pressure is reached during consolidation and filtrate flow declines. Cake is formed until a set-point of low filtrate flow is reached to indicate the end of the cycle. Shortly thereafter, the press is opened mechanically, and the cake is removed (1).

Filter presses are usually installed well-above floor level, so that the cakes can drop onto conveyors or trailers set under the press. Filter presses can be operated at pressures ranging from 5000–20,000 times the force of gravity. In comparison, a solid bowl centrifuge provides forces of 700–3500*g* and a vacuum filter, 1000*g*. Because of these greater pressures, filter presses may provide higher cake solids concentrations (30–50% solids) at lower chemical dosages. In some cases, ash from a downstream incinerator is recycled as biosolids conditioner.

Practice has separated the operation of recessed plate filters into two principal categories: low pressure units and high pressure units, low pressure units operate between 350 and 864 kPa (50–125 psi) as the terminal pressure; high pressure units operate between 1040 and 1730 kPa (150–250 psi). Typically, the low pressure units will terminate at about 691 kPa (100 psi) and the high pressure units at about 1380 kPa (200 psi). There are several ways of maximizing the filtrate removal, including good conditioning and stepping the pressure. Stepping is particularly effective for the high pressure units, using increments of 350–520 kPa (50–75 psi) (2).

The diaphragm (membrane) press is a comparatively newer device, having been commercialized in the United States in the 1980s. It operates during the initial filling period, if the biosolids are properly conditioned, very much like a gravity drainage deck and is able to drain considerable amounts of water at substantially zero headloss across the medium. Overall the diaphragm press operates like the recessed plate press, typically up to pressures between 690 and 1040 kPa (100–150 psi).

The release of water at low pressures helps maintain the integrity of the floc. After water release appears complete following the initial filling period, pumping is stopped and the diaphragm cycle is initiated. The diaphragm pressure is applied, using either air or water on the reverse side of the diaphragm, and pressures up to 1380–1730 kPa (200–250 psi) are applied to the biosolids for additional dewatering. In addition, the confined operation, which follows when the diaphragm pressure is applied effectively, releases substantial additional quantities of water (cake solids will increase 5–8%).

A most significant aspect of the diaphragm press is that its construction and mode of operation allow the use of organic polymers as an alternative to ferric salts and lime conditioning techniques. Although there still is the same tendency to squeeze biosolids into the media itself, the tendency is reduced by the elimination of substantial quantities of water before the start of the squeezing operation.

It should be noted that there has been an evolution in the diaphragm press to a simpler design sometimes known as a diaphragm plate press. In this design, both the cloth and the diaphragm are built into the plate resulting in fewer moving parts, longer cloth life, and much lower O&M (operation and maintenance) costs. Another innovation in pressure filtration is the tubular filter press (5,6), which is still under research and development using bench-scale models.

Based on typical filtration operations, it can be expected that 70–85% of the water will be removed during the low pressure portion of the cycle of the diaphragm press (2). Similar performance can be obtained from a fixed-volume recessed plate press by stepping the pressure at two or three intermediate levels. The diaphragm press, however, usually produces a drier cake than that obtained from the fixed-volume recessed plate. Also, there is a substantially greater uniformity of solids concentration in the cake produced with a diaphragm press. With the low solids feed material continually being supplied to the recessed device, a very low solids cake fraction is produced near the feed point. This problem, of course, is not present in the diaphragm press because the pumping cycle is only the first part of the overall cycle and the diaphragm tends to remove water uniformly. Also the cycle time for a given cake solids concentration is generally less in the diaphragm press.

There appears to be a less frequent need for precoating the diaphragm press than is usually encountered with the fixed-volume device. The implication is that the diaphragm press improves the discharge ability of the cake as a result of the higher cake solids content. Another advantage of the diaphragm press is that the biosolids only need to be pumped in at pressures up to, but rarely exceeding, 865–900 kPa (125–130 psi). The higher pressures during the diaphragm cycle may be supplied by clean water pumps or air pumps, thereby reducing the overall maintenance cost associated with high pressure delivery devices.

While polymers are uniquely successful in conditioning pure WAS and mixed biosolids for dewatering in the diaphragm press, it would be misleading to say that most of the polymer conditioning success has occurred with these units. Actually, with mixed primary and secondary biosolids, pure polymer conditioning has been most successful in low pressure presses operating at 520–1040 kPa (75–150 psi), typically using one or two steps to achieve the ultimate pressure. However the low pressure recessed plate unit does not provide for the final high pressure water removal which the diaphragm press does and this could be the key to better cake discharge from the cloth. The operating sequences for fixed-volume recessed filter and diaphragm presses supplied by different manufacturers are shown in Figs. 3 and 4, respectively.

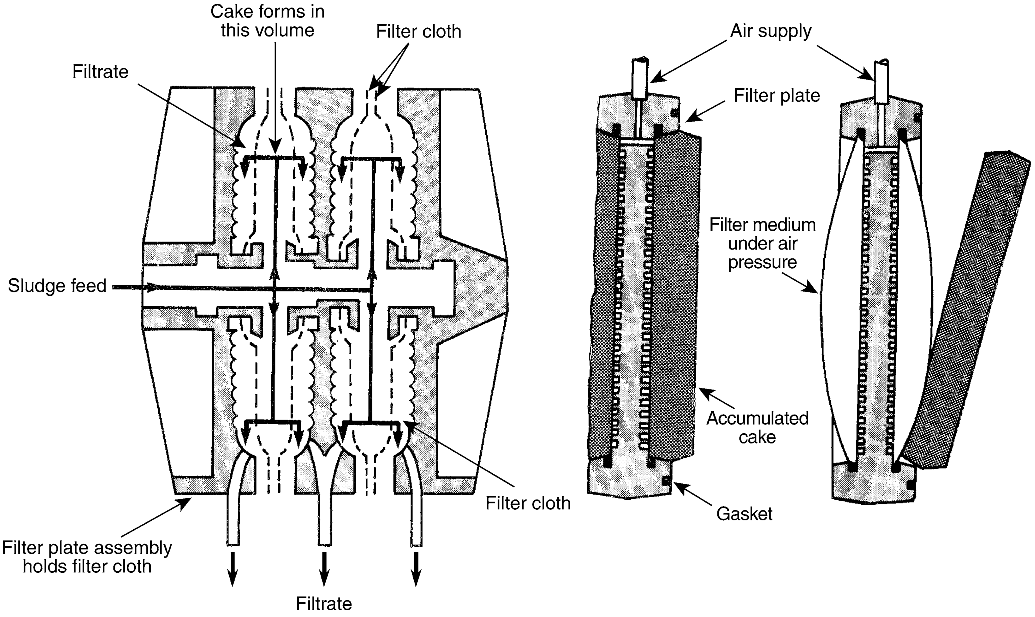

Fig. 3. Filling and cake discharge, fixed-volume recessed plate pressure filter press (*Source:* US EPA).

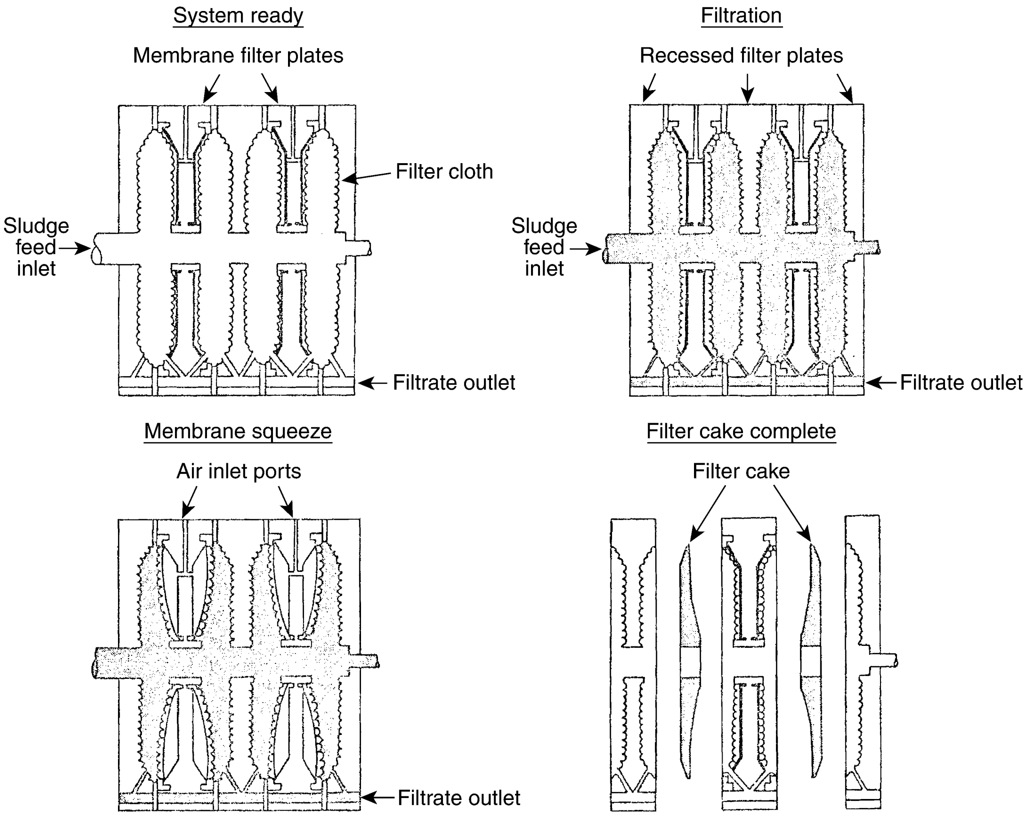

Fig. 4. Filling and cake discharge, diaphragm press (*Source:* US EPA).

Filter press capacity is determined by the number and size of plates and chambers in the press (3). The plates are supported on a structural frame with a shifting mechanism to separate them one at a time. Large presses have automatic plate-shifting systems that press together the plates and filter cloths with a hydraulic ram, sealing the edges of the cloths on the plates, and resisting the filtration pressure developed by the filter feed pump during the filtration process.

Biosolids managers should take into consideration the two types of presses. The conventional press which has a fixed volume removes moisture by adding more solids. The diaphragm press is a variable volume press in which sturdy hollow rubber diaphragm or membrane is positioned behind each filter cloth. Water is pumped to the interior of the diaphragms when the maximum feed pump pressure is reached, expanding the diaphragm and reducing the volume of cake solids. Diaphragm filter presses often result in a cake with higher solids content. Bench testing should be performed on a representative sample of each wastewater treatment plant's biosolids to determine whether a membrane filter offers advantages over a conventional, fixed-volume filter press. An economic analysis should also be conducted to determine whether the additional capital cost of a diaphragm filter press will result in long-term cost savings.

3. APPLICABILITY

Filter presses can be used to dewater most biosolids generated at municipal wastewater treatment plants. Like all dewatering equipment, these filter presses require a capital investment and labor to operate and may not be the most cost effective alternative for wastewater treatment plants operating at less than about 4 MGD (1). The selection of dewatering equipment should be based on the results of a site specific biosolids management plan that identifies processing and end use alternatives and estimates costs. It may be less expensive to haul liquid and pay a processing facility to dewater and process or landfill the dewatered cake. Smaller facilities should also valuate nonmechanical dewatering methods, such as drying beds.

Wastewater plants faced with high end use or disposal costs will benefit from the ability of a filter press to produce the driest cake possible. If the wastewater treatment plant wants to process the cake further, there may be economic advantages to producing the driest cake possible. Incineration or heat drying requires less fuel and may operate at a higher capacity because there is less water to evaporate while alkaline stabilization technologies require less additive, saving on the cost of the additive, storage requirements, and transportation of the final product.

4. ADVANTAGES AND DISADVANTAGES

Recessed-plate filter presses offer several advantages compared to other mechanical dewatering methods, as follows (7,8):

a. High cake solids concentration with associated low biosolids storage, hauling, and disposal costs.
b. Little or no operator attention during dewatering phase of cycle (1–3 h).
c. Cake solids concentration is relatively independent of feed solids concentration.
d. Use of lime as a conditioner stabilizes and disinfects the final product.

The advantages of a diaphragm filter press over conventional recessed-plate filter presses include the following (1):

a. Usually produces a drier cake.
b. Substantially greater uniformity of solids concentration in the cake.
c. Easier to dose polymers as an alternative to ferric salts and lime conditioning techniques.
d. The use of high pressure without having to introduce more liquids reduces the tendency to squeeze biosolids into the filter cloths because substantial quantities of water are eliminated before starting the pressing operation.
e. Removes water uniformly because the pumping cycle is only the first part of the overall cycle.
f. The cycle time for a selected cake solids concentration is usually lower (2).
g. Higher cake solids content improves release of the cake from the filter cloths.
h. Wastewater solids only need to be pumped into the diaphragm filter press at pressures up to 865 kPa (100 psi) reducing maintenance costs (2).

There are also several disadvantages to using recessed-plate filter presses compared with other mechanical dewatering methods, as follows (1,7):

a. Process is mechanically complex.
b. Capital costs are relatively high.
c. Requires special support structure.
d. Requires relatively large area.
e. Filter cloth preparation, cleaning, and cake removal may be operator intensive.
f. Cannot be totally enclosed, leaving operators exposed to odors, gaseous and vaporous sulfur compounds, and ammonia during the cake release phase.
g. When lime and ferric chloride are used in conditioning, they account for a significant portion (15–40%) of the cake solids offsetting the weight reduction of high water removal efficiency.

5. BASIS FOR SYSTEM DESIGN

This section provides an understanding of the following important properties (2):

a. Cake solids concentration.
b. Throughput rate.
c. The recovery fraction or the fraction of those solids delivered to the machine that exit the machine as cake and are not recycled to some other portion of the facility.

The cake solids concentration achievable with particular biosolids will regulate the cost of downstream operations and often determine the need for additional upstream operations such as thickening. There is a relationship between the cake solids concentration and the throughput in that, with filter presses, higher solids are almost always achievable. This is true for any given operating circumstance, if one is willing to increase the cycle time and, therefore, decrease the rate of throughput. The designer's challenge in this regard is to maximize throughput and solids concentration consistency with specific operating conditions.

The third critical design parameter is solids recovery. Systems that do not recover a substantial quantity of solids can experience an increased need for media washing and cause a buildup of fine solids in some process loop, especially one that goes to a thickener or perhaps to the wet end of the plant. Solids losses more than 2–3% of feed solids are usually traced to torn media or biosolids adhering to the media on discharge and washed off to be recycled. Sometimes this buildup of fines can lead to higher effluent suspended solids concentrations. In any event, recovery in excess of 95% is an important design objective of a system and is necessary to prevent both the excessive recycle of solids and the possible impact on some aspects of the wastewater treatment plant's operation. In this regard, filter presses generally are superior, with solids recovery typically more than 98%.

Filter presses are sized based on the volume of solids to be dewatered. To determine the number and size of presses for a project, the following information must be determined (1):

a. Amount of primary biosolids that will be flowing through the plant per day.
b. Amount of secondary biosolids produced per day.
c. Amount of tertiary biosolids produced per day.
d. Volume of thickened biosolids to be dewatered per day.
e. Seasonal variation in biosolids production.
f. Range of solids concentration in the feed biosolids.
g. Future increases in biosolids.
h. Changes in biosolids quality or quantity from industrial sewer users or in-plant process changes.

An effective biosolids management plan will include the above information. It is important to design for excess capacity to ensure that the anticipated amount of incoming biosolids can be easily dewatered during operating hours. Allowing for excess capacity ensures that a plant will not experience a build-up of biosolids if one unit is out of service. If only one unit is required, the plant should have an alternate program to remove solids in liquid form for transport to an alternate processing site.

Pilot testing by the vendor offers the best way to obtain data on the important design aspects (2). Pressure is determined by filter feed-pump output. Presses are usually designed to operate at 689.5 kPa (100 psi) or 1551 kPa (225 psi) terminal pressure. Progressive-cavity and piston-membrane pumps have been used as filter-feed pumps in these systems. A progressive-cavity pump must have variable-speed and high turndown-ratio capabilities to meet low flow requirements at the end of the cycle (3). A piston-membrane pump automatically compensates for increasing pressure by bypassing hydraulic fluid and reducing the pump's stroke volume.

Filter press installations are mechanically complex (3). System components may include conditioning tanks, mixers, multiple chemical-feed systems, feed pumps and a filtrate removal system. Ferric chloride requires corrosion-resistant facilities and extreme caution in handling. The press has a hydraulic power-pack system and other mechanical accessories for plate shifting and washing, as well as drip trays. System components are generally reliable, but require routine inspection and lubrication.

Most buildings must be custom designed to accommodate a filter press. The cake is released to fall into a bin below the floor of the press and must be moved to a truck or roll-off container. Because filter presses operate in a batch mode, the system may require a liquid storage tank. The operator may want to remove biosolids from a digester or settling tank in small quantities every 15 min rather than in large quantities every several hours. A wide variety of filter cloth material is available. The manufacturer should test to determine the best cloth for each facility. Selecting the correct filter cloth will improve release of cake, minimize cleaning requirements and maximize service life.

6. EVALUATION OF DESIGN PARAMETERS

Treatability testing is used to evaluate design parameters and the potential effectiveness of the filter press. This testing may begin at the bench-scale level and proceed to pilot-scale or full-scale testing. However, if pilot-scale testing is not feasible, the design can be developed from the bench-scale data (9,10).

6.1. Types of Tests

The types of tests that can be conducted include basic filterability tests and tests to optimize chemical conditioning (10).

a. **Basic filterability tests.** Basic filterability testing evaluates the filtering properties of the biosolids and determines the ease of separating the water phase from the solid phase (2,7). Two basic parameters that can be used to provide design information on final solids concentration are specific resistance and capillary suction time (CST). The specific resistance testing can be used as a basic guide in estimating the solids yield and cake solids. CST tests can also be used to evaluate whether the biosolids can be easily dewatered; however, they are primarily used to evaluate the effectiveness of biosolids conditioning.
b. **Conditioning tests.** Optimizing chemical dosages is not only important to the dryness of the cake, but it also affects the solids capture rate and solids disposal costs. Several types of tests can evaluate the effectiveness of a single conditioning chemical or group of conditioning chemicals. Standard test procedures include jar tests, CST tests, Buchner funnel tests, and pilot-scale and on-line testing. Although chemical dosages should be initially evaluated, they should also be reevaluated periodically because of changes in biosolids characteristics.

6.2. Test Procedures

Test procedures that can be used for basic filterability testing and conditioning testing include jar tests, CST tests, specific resistivity tests, and pilot and on-line tests (10).

a. **Jar testing.** Jar testing, the simplest type of conditioning testing, is often used for the preliminary evaluation of the type and estimated quantity of conditioners required. Jar testing involves visually observing the size of biosolids floc produced when various types and quantities of different types or combinations of conditioning chemicals are mixed with samples. This type of testing can be used to screen or eliminate different types of chemical conditioners and determine the effects of different dosages of a specific conditioner. A description of the jar testing procedure is outlined in the United States Environmental Protection Agency (US EPA) Design Manual-Dewatering Municipal Wastewater Sludge (2).
b. **Capillary suction time (CST) testing.** The CST test involves measuring the time to move a volume of filtrate over a specified distance as a result of the capillary suction pressure of dry filter paper. The CST test provides information regarding the ease of separating the water portion from the solids portion of biosolids. This type of testing is most effectively used during the selection of the optimum conditioner dosages during on-line tests. The CST is typically defined in units of time (s). A detailed theoretical description of this method and its procedures are presented in the US EPA Design Manual-Dewatering Municipal Wastewater Sludge (2).
c. **Specific resistance testing.** Specific resistance testing has been widely used and investigated as a way to evaluate the effectiveness of filterability. Specific resistance is typically defined in units of tetrameters (10^{12} m) per kilogram (Tm/kg). A lower specific resistance indicates increased dewaterability. This testing can be done by calculating the specific resistance from Buchner funnel testing or by measuring the specific resistance directly with

specific resistance test meters. The Buchner funnel test is a method commonly used for predicting the specific resistance of biosolids. A detailed theoretical description of this testing and its procedures are provided in the US EPA Design Manual-Dewatering Municipal Wastewater Sludge (2).

d. **Pilot-scale and on-line testing.** Based on the results of jar tests or the other tests previously described, pilot-scale testing or on-line testing can be done to evaluate different conditioners and to determine their optimum dosage based on actual thickening or dewatering performance. During pilot-scale and on-line tests, actual samples of the raw biosolids feed, conditioned biosolids, thickened or filter cake discharge, and filtrate or supernatant are collected and analyzed. On the basis of these test results, an economic analysis may also be conducted as part of the final evaluation of the optimum dosage. In addition to evaluating and optimizing conditioning agents, pilot-scale and on-line testing can be used to determine filter press operating conditions, such as optimal filtration cycle times and pressures (i.e., feed, compression, extraction), required filtration area, and the need for filter media precoat and filter aids. Pilot-scale testing is commercially available from several filter or filter press manufacturers that have bench-scale equipment (i.e., cylinders or plate unit) and trailer mounted equipment which can simulate actual operating conditions. In addition to pilot-scale testing, on-line testing to verify optimal operating conditions should be done after the filter press is installed. During this testing, conditioning dosages may be further optimized, and actual dewatering operation conditions (such as cycle times) and equipment selections (such as the filter media or the need for precoating or filter media washing) may be further refined.

7. DESIGN PROCEDURES

This section contains a review of the methods employed for predicting solids concentration, throughput, and recovery based on the rather simple and straightforward laboratory tests discussed in the previous section. Pilot operations, if feasible, offer the best way of obtaining data on all three of the important design aspects. However, pilot operations are often not possible, in which case it becomes necessary to design from bench-scale information.

Specific resistance and CST are used to develop design information on throughput and final solids concentration. Most of the relationships discussed below utilize the specific resistance test as the basic guide in estimating yield and cake solids. Yet, the CST is also a useful test, and this section contains several references to its use.

There are some significant dimensional considerations which must be discussed for a full understanding of specific resistance. Christensen (11) has summarized typical values of specific resistance for water residuals and wastewater biosolids and commented on the disparity in the use of units to describe specific resistance. He points out that s^2/g probably is an incorrect unit assignment because of the manner in which these units were first employed. Gale (12) has pointed out that s^2/g as a unit describing specific resistance was a result of improperly using g/cm^2 for pressure difference across the filter cake. It has been suggested that m/kg is a more satisfactory unit.

Christensen suggests the use of Tm/kg as the best possible unit. Christensen has also noted that proper conditioning, generally speaking, changes the specific resistance by a factor of 10^2–10^3. Raw wastewater biosolids have specific resistance values of 10–100 Tm/kg (1.5×10^{13}–15×10^{13} ft/lb). Adequately conditioned solids have specific resistance values of about 1 Tm/kg, and well conditioned solids have specific resistance values on the order of 0.1 Tm/kg. The conversion factors to go from s^2/g to cm/g and m/kg are 9.81×10^2 and 9.81×10^3, respectively.

The CST test provides a substantial amount of information about the ease in separating the water portion from the organic solids portion of biosolids. For example, unconditioned WAS has a CST of 100–200 s. For a filter press to function, dewater, and release the WAS cake, a CST of 10 s or less is required.

The following series of relationships show the development of the significant equations that govern flow through porous media and, hence, filtration phenomena. In the 1800s, Poiseuelle described the velocity in a circular capillary tube as (2):

$$U = \frac{d^2 g}{32\,\mu} \times \frac{Pg}{L} \tag{1}$$

where U is the linear velocity; d is the diameter of capillary; P is the pressure differential; μ is the viscosity of the liquid; L is the length of capillary; and g is the gravitational constant.

D'Arcy also in mid-1800s showed that $(d^2g/32\mu)$ is a constant by noting that the flow through sand beds may be described by:

$$U = K_1\left(\frac{P}{\mu L}\right) \tag{2}$$

These equations were modified by Kozeny (13), who introduced porosity and specific surface in the equation:

$$U = \left[\frac{\varepsilon^3}{(1-\varepsilon)^2}\right]\left[\frac{1}{K\mu S_o^2}\right]\left[\frac{Pg}{L}\right] \tag{3}$$

where ε is the porosity; S_o is the specific surface; K is the a constant equal to 5. Hence, with filtration:

$$U = \left(\frac{1}{A}\right)\left(\frac{dv}{d\theta}\right) \tag{4}$$

And

$$U = K_1\left(\frac{P}{\mu L}\right) \tag{5}$$

where

$$K_1 = \frac{g\varepsilon^3}{5S_o^2(1-\varepsilon)^2} \tag{6}$$

$dv/d\theta$ is the rate of flow of liquid across cake. This form leads to the conventional expression:

$$\frac{dv}{d\theta} = \frac{PA}{\mu RL} \tag{7}$$

where L is the cake thickness; R is the specific resistance (s^2/g); A is the area of cake; but, LA is the cake volume and also, cake volume = $v\ V$ where v is the volume of solids deposited per unit of filtrate; and V is the volume of filtrate.

Hence $LA = vV$.

$$L = \frac{vV}{A}$$

and then by substitution,

$$\frac{dv}{d\theta}=\frac{PA^2}{\mu RvV} \tag{8}$$

Carman (14), noting that R must include all resistance, developed an equation with two terms—one for the cake and one for the media:

$$\frac{dv}{d\theta}=\frac{PA^2}{\mu(rvV+R_mA)} \tag{9}$$

where r is the specific resistance of cake; v is the mL of cake deposited by 1 mL of filtrate; and R_m is the initial resistance of 1 cm^2 of filtering surface.

For compressible cake, vV becomes Vc where c is the weight of dry solids per unit volume in the unfiltered slurry. The general equation then becomes:

$$\frac{dv}{d\theta}=\frac{PA^2}{\mu(rcV+R_mA)} \tag{10}$$

Integrating the expressions and neglecting the resistance of the media, the time for filtration becomes:

$$\theta=\frac{\mu rcV^2}{2PA^2} \tag{11}$$

If θ/V is plotted against V, a straight line is obtained where V slope, b, is:

$$b=\frac{\mu rc}{2PA^2} \tag{12}$$

Therefore, the specific resistance may be calculated from Buchner funnel test data where,

$$r=\frac{2bPA^2}{\mu c} \tag{13}$$

Specific resistance has been used in calculating yields from pressure filters. Coackley (15) reported a procedure in 1957. Mininni, Spinosa and Misiti (16) have presented a procedure for predicting the filtrate flow rate and cake concentrations for fixed-volume pressure filter filtration. These workers observed that φ, the filtrate flow rate or flux, after the initial period of drainage or while the cake is being formed, is described by the expression:

$$\varphi = at^b \tag{14}$$

where φ is the filtrate flow rate or flux; t is the time; *a and*, b are the coefficients. The coefficients can be determined if the specific resistance, initial solids concentration, filtrate viscosity and maximum operating pressures are known. The final cake concentration then can be calculated by making a material balance, assuming that the dry solids density, the filtering time, the conditioner dosage, the slurry concentration, filter press chamber volume, and filtration surface areas are all known. The reported agreement between predicted and actual values is excellent (2).

Wilhelm (17) obtained the following expression from classical filtration theory:

$$\log\theta = \log\left[KA^2\left(\frac{S_c}{c}\right)^2\right] + \log(\ell\rho_f)^2 \tag{15}$$

where θ is the filtration time (min); A is the filtration area (cm^2); S_c is the cake solids concentration by weight fraction; c is the feed solids concentration (g/cm^3); ℓ is the cake thickness (cm); ρ_f is the feed density (g/cm^3).

Wilhelm's procedure provides excellent correlation when the cycle time is plotted against his correlating factor:

$$KA^2\frac{S_c^2}{c^2}$$

To obtain good replication on different runs with the same biosolids requires the solution of similar ranges of pressure to obtain "K" values. The role of the filter media and the relationship between the character of the material being filtered and the media has been described in a study by Christensen and Sipe (18). They developed the following equation which, like the Carman equations, separates the resistance associated with the cake itself from that associated with the medium. The equation is:

$$\frac{t}{V} = \frac{\mu Rc}{2P_tA^2}\left(V + \frac{\mu R_m}{P_tA}\right) \tag{16}$$

where t is the time; V is the filtrate volume; μ is the absolute viscosity; R is the specific resistance (in this case m/kg); c is the mass of cake deposited per unit volume of filtrate; P_t is the total pressure drop across cake and medium; A is the filtration area; R_m is the resistance of the medium.

If the medium resistance is negligible, the preceding equation can be rearranged to a form similar to Carman's equations:

$$\log t = \frac{\mu Rc}{2P_cA^2} + 2\log v \tag{17}$$

where P_c is the pressure drop across cake. The authors suggest that the rearrangement of the equation offers the second way to plot filtration data, i.e., log t vs log v. When this is done, according to the equation, the data should plot as a straight line with the slope of two. The intercept at a convenient point, such as volume = 1, can be used to calculate the specific resistance, since all of the other factors in the first term of the equation are known. The authors also point to many significant deviations when the slope is equal to two; one is that the equation can be written as $t = KV^n$, where n varies. In the t/V vs V plot (which will be linear only if n is equal to two when the data is approximated by a straight line), the intercept of that line will be negative when n is more than two and positive when n is less than two. The t/V intercept is proportional to the medium resistance.

Constructing a log t vs log V plot of the data from a filtration experiment with significant medium resistance is equivalent to satisfying the equation: $\log t = \log(K_1V^2 + K_2V)$

where K_1 represents the constant items in the initial term of the first of equation of Christensen and Sipe and where K_2 represents the constants in the second term.

Noting a consistent deviation from theoretical practice, Notebaert et al. (19) have proposed a modification of the standard cake filtration model to account for the deviations. The significant findings of Notebaert et al. (19) were summarized by Christensen and Sipe (18) as follows:

a. The assumption that the medium does not become fouled or clogged is not realistic.
b. If the particles in the biosolids are of the same order as the pores in the filter medium, the medium will clog. If the particles are a great deal larger than the pores in the medium, clogging will still occur but it will occur over a much longer period.
c. When the medium is clogged, resistance will be high at the beginning of the filtration cycle while the medium is clogged, but will increase slowly afterwards. Therefore, the average specific resistance will be decreasing throughout filtration. If the cake becomes clogged, the clogging will continue throughout filtration with a continuous increase in the average specific resistance.
d. The slope of the log t vs log V plot is indicative of the physical processes described. If the medium is clogging, the slope will be less than two. If the cake is clogging, the slope will be more than two.

Selection of the optimum filter media, based on the manufacturer's specified characteristics of the media (which will include data such as air flow, the weave, the fabric, and so on) is not yet possible. However, the Metropolitan Waste Control Commission of the Twin Cities (20) has carried out a detailed and comprehensive study on pressure filtration. When wastewater, without solids, comes in contact with the filter medium, the media resistance will increase. This is probably because of bacterial growth, as the presence of chlorine decreases the rate at which resistance increases. Pressure and the impact of pressure on the fibers themselves increase the extent and rate of blinding. The workers observed that polypropylene and nylon are the two most commonly used materials for filter cloth. The authors, in their literature review, pointed to Purchas' work. He tried to relate filtrate clarity, resistance to flow, cake solids, ease of discharge, cloth life, and tendency to blind to media characteristics. They also noted that criteria set forth by Warring might be the best and most reliable guide for establishing a good model. These criteria are:

a. How small a particle can the media retain?
b. What is its resistance to flow?
c. What is the relationship between buildup of particulates in the medium to the rate of flow?

Finally, these workers concluded from their own studies that the resistance of the filter cloth increased markedly with use. Periodic washing with water or with acid reduced media resistance. The effect of media resistance on press operation was found to be very significant. Cake solids decreased from 60 to 33% on a full-scale press because of increased media resistance. Filtration rates, filtration yield, mass of dry solids deposited, and cake percent solids all decreased because of increases in media resistance. Any model of a pressure filtration process must include a term for media resistance. Typical loading rates and expected performance from pressure filtration of various biosolids are shown in Table 1 (2).

Table 1
Typical Design Parameters and Performance of Pressure Filtration

Biosolids type	Conditioning	Solids to pressure filter (%)	Typical cycle length (h)	Solids filter cake (%)
Primary	5% $FeCl_3$, 10% Lime	5	2	45
	100% Ash		1.5	50
Primary + $FeCl_3$	10% Lime	4[a]	4	40
Primary + 2 stage high lime	None	7.5	1.5	50
Primary + WAS	5% $FeCl_3$, 10% Lime	8[a]	2.5	45
	150% Ash		2	50
Primary + (WAS + $FeCl_3$)	5% $FeCl_3$, 10% Lime	8[a]	3	45
(Primary + $FeCl_3$) + WAS	10% Lime	3.5[a]	4	40
WAS	7.5% $FeCl_3$, 15% Lime	5[a]	2.5	45
	250% Ash		2	50
WAS + $FeCl_3$	5% $FeCl_3$, 10% Lime	5[a]	3.5	45
Digested primary	6% $FeCl_3$, 30% Lime	8	2	40
Digested primary + WAS	5% $FeCl_3$, 10% Lime	6–8[a]	2	45
	100% Ash		1.5	50
Digested primary + (WAS + $FeCl_3$)	5% $FeCl_3$, 10% Lime	6–8[a]	3	40
Tertiary alum	10% Lime	4[a]	6	35
Tertiary low lime	None	8[a]	1.5	55

Source: US EPA.
[a] Thickening used to achieve this solids concentration.

8. SUPPORT EQUIPMENT AND PROCESSES

8.1. Biosolids Conditioning Process

Biosolids must maintain some structural integrity during the pressing period, because a massive structure will prevent the movement of water through the filter cake to the discharge or filtrate side. To this end, one of the essential parts of a filter press system is the biosolids conditioning subsystem. For recessed volume filter presses, the most common conditioning technique for digested and WAS—and probably the most common for primary biosolids—is the addition of iron salts and lime (21,22). The average quantity required is on the order of 5% ferric chloride and 20% lime, though values as low as 3% ferric chloride and 10% lime and as high as 10% ferric chloride and 40% lime have been reported and are sometimes required (2).

In general, the ferric chloride requirement is a function, at least in part, of the biosolids alkalinity. The role of lime in biosolids conditioning has been discussed extensively by Webb (23) and Sontheimer (24), but is still somewhat unclear. Lime's solubility above pH 11.0 or 12.0 is only on the order of 1 g/L. As a result, much of the lime

Table 2
Comparison of Iron Conditioners With and Without Lime

Total sludge solids[a] (%)	Iron conditioner	Iron dose (%)	CST after iron addition (*s*)	Lime dose CaO (%)	Specific resistance after iron and lime addition (10^{11} m/kg)
5.5	$FeSO_4 \cdot 7H_2O$	1.72	208	15	14
5.5	$FeCl_2 \cdot 4H_2O$	1.72	157	15	7.9
5.5	$Fe_2(SO_4)_3 \cdot 6H_2O$	1.72	41	15	5
5.5	$FeCl_3 \cdot 6H_2O$	1.71	26	15	2.6
5.5	$FeSO_4 \cdot 7H_2O$	3.44	180	30	6
5.5	$FeCl_2 \cdot 4H_2O$	3.44	139	30	2.9
5.5	$Fe_2(SO_4)_3 \cdot 6H_2O$	3.44	27	30	2.3
5.5	$FeCl_3 \cdot 6H_2O$	3.44	19	30	1.2
7	$FeSO_4 \cdot 7H_2O$	3.44	480	20	11
7	$FeCl_2 \cdot 4H_2O$	3.44	–	20	5.6
7	$Fe_2(SO_4)_3 \cdot 6H_2O$	3.44	117	20	5.3
7	$FeCl_3 \cdot 6H_2O$	3.44	58	20	1.8

Source: US EPA.

[a] The activated sludge employed was a raw mixed sludge approx 50% primary and 50% waste-activated on a dry solids basis.

must exist as partially hydrated calcium hydroxide, which probably acts structurally to provide channels for water to move through to the filtrate side.

In studies with four iron salt conditioners and lime, Christensen and Stule (25) obtained both CST data and specific resistance data. The results are shown in Table 2 together with a correlation between these two for the particular biosolids under study. This data is presented in Fig. 5 and shows a remarkably good correlation. Briefly, the ferric conditioners performed the best and the chloride form was more effective than sulfate ore which appeared to produce a slightly poorer floc and a more poorly conditioned cake.

The role of calcium was also studied extensively, and it was concluded that calcium is involved in a chemical link with the iron floc. However, calcium chloride was used and the pH was raised with sodium hydroxide to obtain a somewhat synthetic situation. Extensive reports exist which relate the necessity to clean calcium hydroxide scale off both media and plates, indicating that considerable quantities of calcium hydroxide exist when the biosolids are conditioned with lime. Two other important observations are that aging the biosolids can as much as double the specific resistance in one hour. The actual impact of aging depends on the flocculants used. Experience indicates that prolonged agitation and tank storage time associated with batch conditioning can result in a feed of varying and deteriorating dewaterability. For this reason, conditioning processes are now frequently designed to provide "in-line" conditioning. This can be

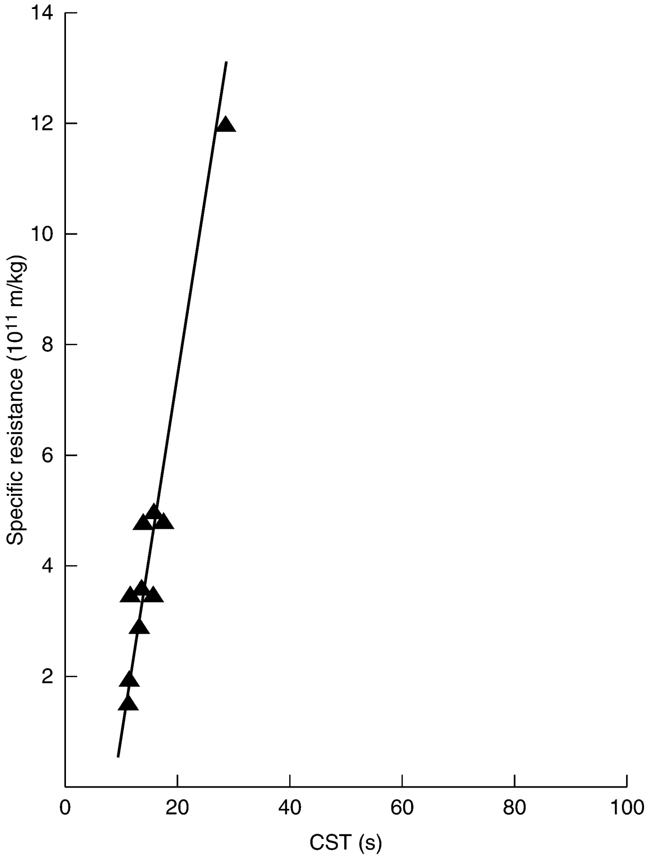

Fig. 5. Specific resistance vs capillary suction time (*Source:* US EPA).

accomplished by either the continuous pumping of biosolids into a small tank and addition of chemicals, or directly injecting conditioning chemicals into the biosolids on the way to the filter. In-line conditioning diminishes the deleterious effects of storage and prolonged agitation. Figure 6 shows a schematic for in-line conditioning. Also, it is clear that the pH has a substantial effect. The specific resistance dropped from 1.6×10^{12} m/kg at a pH value of around 11.3 to about 0.2×10^{12} m/kg at a pH of 12.45. This clearly is a pH related phenomenon and not tied into any physical property of calcium because no precipitation of calcium hydroxide was observed.

The use of polymers for biosolids conditioning is expanding. Polymers can produce very nearly similar cake solids and do not result in a 15–30% increase in cake weight and volume. Dosing procedure, flocculation requirements, and filter press pressure-time relationships necessary to optimize polymer dosage and cake release are site specific. About 75–80% of the conversion to polymer trials was successful. Polymer costs are 30–70% of ferric and lime costs.

8.2. Feed Pump System

One major problem with pressure filters has been the need to design a system that will pump from 30 to 2000 gpm (1.9–126 L/s) of a viscous, abrasive slurry at pressures

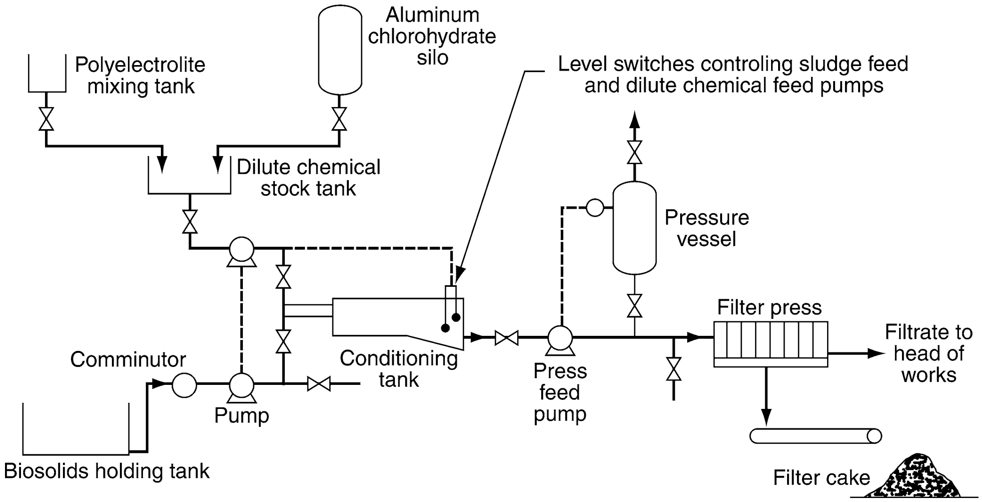

Fig. 6. Schematic of an in-line conditioning system for recessed plate pressure filter press (*Source:* US EPA).

of 40–225 psi (276–1551 kN/m^2). Ideally, the feed system should inject conditioned biosolids into the chamber as rapidly–as possible but slowly enough to permit sufficiently prompt formation of a uniform and thick enough cake to prevent any incursion of biosolids particles into the filter cloth. Imbalance of the biosolids feed and cake formation rates can result in non uniform, high resistance cake, or in cloth blinding and/or initial poor filtrate quality. If a nonuniform cake is formed or excessive fines migrate, then a long filter cycle or an inordinate amount of cloth plugging will result (26).

The filter feed methods used for some pressure filters involve a combination of pumps and pressure vessels (27). These combinations are used to obtain a high initial feed rate of approximately 2000 gpm (126 L/s) through the pressure vessel, followed by the use of reciprocating ram high pressure pumps to pump at a pressure of 225 psi (1551 kN/m^2) at feed rates of 100–200 gpm (6.3–12.6 L/s). In some cases, a combination of progressive cavity pumps and pressure vessels is used for the lower pressure, high-rate chamber filling phase.

8.3. Cloth Washing and Cleaning

Because recessed plate pressure filters operate at high pressures and because many units use lime for conditioning, the designer must assume that cloths will require routine washing with high pressure water, as well as periodic washing with acid. Practices vary according to the particular biosolids and proprietary process. Designers should ask for recommendations from equipment manufacturers (28–33) on frequency of washing. Occasionally it might be necessary to wash the cloths in place. When this is done the cloths or media should be pulled square and free of any creases. A hand-held, high pressure, single jet (about 750 psi) is usually effective for cleaning cloth of the media. A plastic cover draped over the filter will be needed to confine spray during the cleaning cycle. Mechanized washing arrangements are available for some filters. Where acid washing is provided, a recirculating system provides both a scrubbing and acid effect as opposed to merely soaking the media in acid.

8.4. Dewatered Cake Breakers

Design of suitable breakers is a function of the structural properties of the dewatered cake. Pressure filter cake is usually friable enough that use of breaker wires, bars, or cables beneath the filter will be sufficient. If, however, polyelectrolyte conditioning is contemplated, consideration should be given to the resulting changes in cake structure.

9. OPERATION, MAINTENANCE, AND PERFORMANCE CHARACTERISTICS

Unless the inorganic content of the feed solids is high, conditioning chemicals are required for successful filter press dewatering (3). In the past, filter presses relied on lime and ferric chloride for conditioning. While these chemicals typically produced a dewatered cake with more than 40% solids content, they increased the mass to be stored, transported, used or disposed. Lime is also associated with ammonia releases which must be considered in overall facility design, including ventilation, and odor control requirements.

Operating a filter press manually in a small plant is simple (3). Batches of solids are preconditioned and fed to the press. Monitoring is not required during the filtration cycle. If cake release is good, it will drop cleanly from the cloth when the plates are shifted. In larger units, plate shifting is automatic, whereas smaller units use a power-assisted plate shifter.

Facilities with multiple presses need a fully automatic system for efficient operation to maintain proper chemical dosages, open and close the press, and blow out the core at the end of the cycle. Even in an automated system, the plate-shifting step must be initiated manually so the facility can prepare to receive the cake drop.

Filter cloths require periodic washing. Larger presses have automatic washers that require a high-pressure pumping system to supply spray water. In some installations, the press can be filled with an acid cleaning solution to remove scale deposits when lime is used for conditioning. Acid washing may reduce filter cloth life and replacing filter cloth is labor-intensive. Filter cloth life depends on the material, solids type, conditioning, and washing frequency.

The use of polymer in variable speed presses requires automation to control the dose as polymer dose is related to the volume of filtrate exiting the press. High cake solids are possible, but cycle times are long and the cake often sticks to the cloth, requiring assistance for its removal. Residual solids often remain on the cloth, reducing solids capture and requiring more frequent cleaning. Polymer conditioning is most successful in diaphragm systems because dosing is not as complicated. Filtration cycle times with lime and ferric chloride conditioning range between 1 and 3 h. With polymer conditioning, filtration cycles might exceed 3 h and tend to dewater the core, making core blowout ineffective. An advantage of polymer conditioning is that it produces cake with fewer inert solids, enabling more solids to be processed per cycle than with lime and ferric chloride. The cake contains more volatile solids than when processed with inorganic conditioning and therefore can be disposed of by incineration or other type of thermal processing.

The degree of operator activity associated with filter presses is similar to that of belt presses. Although the press operates unattended during filtration, the system uses a batch process that requires regular operator attention to fill and unload the press (3). When filtration is complete, compressed air should be used to blow out the core since it is filled with partially dewatered cake at the end of the filtration cycle.

Record-keeping is essential for all operations that require conditioning (2). The operator must keep a log of the lime and ferric or polymer dosage required to reach a given degree of cake solids with a particular blend or type of solids. It is also helpful to keep track of pressing time and filtrate quality to gauge performance of the filter cloths.

9.1. Control of Machine Variables

In a conventional filter press, the operator controls the following variables (2):

a. Pressure of the feed biosolids.
b. The rate at which the pressure is applied and the pacing of flow to the filter press.
c. The overall filtration time, including such variables as the time at each pressure level in multiple pressure level operations.
d. The use of precoat or body feed and the amount of material used.
e. Conditioning chemicals including type, dosage, location, mixing efficiency, and flocculation efficiency.
f. Cloth washing frequency.
g. The nature of the filter media used.

A similar set of machine variables exists for the diaphragm filter press. They are as follows (2):

a. Pressure of the feed biosolids and the rate at which the feed is added to the machine.
b. Filtration time.
c. Diaphragm pressure.
d. Diaphragm squeezing time.
e. Rate at which the diaphragm pressure is increased.
f. Conditioning chemicals including type, dosage, point of addition, mixing efficiency, and flocculation efficiency.
g. Filter media used.
h. Cloth washing frequency.

Changes in these parameters are predictable up to a point, and mechanisms exist to evaluate the effect of varying each one for optimizing the system. Precoat generally does not need to be used when inorganic conditioning chemicals, particularly ferric chloride and lime, are used. Heavy doses of organic polyelectrolyte may also preclude the use of body feed or precoat. Precoat is normally used in cases where the particle size is extremely small or considerable variability in filterability and substantial loss of fine solids to and through the filter media are anticipated. A final decision about the need for precoating may require lab or field experimentation with the specific biosolids.

When substantial quantities of lime are used, cloth washing may require both an acid and a water wash. Therefore, a medium is needed that is resistant to both acid and alkaline environments. In those instances where polyelectrolytes are used, the washing operation normally is accomplished with only clean water since the biosolids imbedded in the media are backflushed to the waste.

9.2. Control of Process Variables

Few process variables, as opposed to machine variables, are likely to be controllable by the operator. Process variables include (2):

a. **The type of biosolids to be dewatered.** Raw or digested primary biosolids, WAS, trickling filter biosolids, RBC biosolids, or mixtures thereof have varying effects on the dewatering process.
b. **The age or the freshness of biosolids.** Conditioning, particularly conditioning with polyelectrolytes, is much more dependable and reproducible when the biosolids are fresh. The specific resistance increases with time. Therefore, it is desirable to dewater biosolids in as fresh a condition as possible.
c. **Prior chemical conditioning.** Prior chemical conditioning tends to confound the use of chemical conditioning at the dewatering device. This is particularly true when polymers are used and if polymers have already been used somewhere upstream from the dewatering system. If this condition exists, the best remedy is to use a small quantity of the polymer of the opposite or neutral charge, followed by the normal dose of the polymer usually employed. Establishing charge reversal with the polymer of the opposite charge eliminates the confounding effects of the old, partially degraded polymer on biosolids surface.
d. **The solids concentration achievable in the final clarifier or in subsequent thickening operations.** Generally speaking, it is desirable to send to the dewatering device feed biosolids with the highest possible solids content.
e. **Solids capture.** If the cloth is unbroken and cake cleanly discharged, suspended solids recovery is about 99%. When the cloth is washed, the effluent solids are somewhat higher.
f. **Cake concentration.** The cake concentration must be sufficiently high to readily discharge from the cloth. Variables affecting cake concentration have been reviewed earlier.
g. **Throughput rate.** The throughput will be dependent on the water release characteristics of biosolids, type and amount of chemicals, and the desired minimum cake solids.
h. **Conditions under which the biosolids were produced.** The filterability of biosolids, particularly WAS, is strongly dependent on the conditions under which the biosolids were produced. This consideration probably applies to those municipal wastewater treatment plants receiving substantial quantities of high carbohydrate industrial wastes that may produce, on occasion, a nitrogen deficient situation in the activated sludge portion of the plant. However, nitrogen deficient activated sludge has a considerably higher specific resistance when untreated than activated sludge grown under nitrogen enriched conditions. In addition, the final specific resistance after chemical conditioning is not as good as that achieved with activated sludge grown under excess nitrogen conditions. The conditioned specific resistance of the nitrogen-poor biosolids generally runs two to three times that of the activated sludge grown under high nitrogen conditions when properly chemically treated (34).

9.3. Control Considerations

1. **Physical control.** Instrumentation is usually minimal; however, it is possible to completely automate the operation of the filter press if desired. Pressure gauges should be provided to monitor the feed pressures and the filtrate flow must be monitored either visually or with a flow indicator.
2. **Moisture control.** If the filter press is operated as recommended with sufficient washing and air drying time between cycles, the cake should have the highest possible solids content. It should discharge from the press with a minimum of debris left behind. Discharge of a wet cake can lead to dirty cloths on the lower stile faces making it difficult to obtain a good seal on this gasket area when closing the press.
3. **Filtrate flow.** It is usually possible to develop an excellent relationship between filtrate flow rate (which decreases as the cycle progresses) and cake moisture for a given

Table 3
Common Design Shortcomings

Shortcoming	Solution
Gravimetric ash feeders installed—bulking problems with ash	Install volumetric feeders
Cake transport system inadequate (screw conveyors plug; belt conveyor limited to 15° slope)	Install heavy duty flight conveyor
Mechanical ash conveyor installed—noise and maintenance problems	Install pneumatic ash conveying system
Improper media specified—poor cake discharge, difficult to clean	Change media—usually monofilament, relatively coarse media are used on municipal biosolids

Source: US EPA.

sludge. That is, for any given filtrate flow rate, a corresponding filter cake concentration can be expected.

4. **Precoat.** Whether or not to precoat is an operational question. The precoat is the placement of an initial coating on the filter cloth prior to application of biosolids. The precoat acts as an additional filtration membrane and also aids in a clean removal of biosolids from the cloth. If the investment in a precoat system has been made, its use should reduce manpower requirements for media cleaning and may provide better performance.
5. **Cloth characteristics.** If the press is operated as recommended, but performance of the cloth is unsatisfactory a different type of cloth may give better results. If this happens, other cloth types can be tried until the desired performance is obtained. The addition of precoat may also aid in performance.
6. **Sampling.** Sampling should be performed as required for process control. Samples may be obtained through valves provided in the respective piping or directly from the process. If sampling points are not provided, they should be installed to facilitate operation and control of the process. Samples should be analyzed according to procedures specified in Standard Methods.
7. **Control of common shortcomings.** Solution to common design shortcomings are listed in Table 3.
8. **Evaluation of performance.**
   a. Check the filter cake solids concentrations against typical values in Table 1.
   b. Check the length of filter cycle against Table 1.
   c. Check the quality of the filtrate. With proper operation and conditioning, the suspended solids concentration should be less than 100 mg/L.

Refer to Troubleshooting Guide shown in Tables 4 if unusual values are found.

10. SURVEY OF FILTER PRESSES

In order to provide detailed information on the operating experience of filter presses, a survey was made of 50 municipal wastewater treatment plants by Terraqua Corp. of Hunt Valley, MD (35) for the City of Baltimore. The plants ranged in size from 0.66 to 265 m^3/min (0.25–100 MGD).

Table 5 summarizes general information on the filter press installations. Of the 50 plants, 42 were dewatering anaerobically digested or raw biosolids: 21 of each biosolids type. One plant was elutriating anaerobically digested biosolids, five were processing

Table 4
Trouble Shooting Guide

Indicators/observations	Probable cause	Check or monitor	Solutions
Plates fail to seal	Poor alignment	Alignment	Realign plates
	Inadequate shimming	Stay bosses	Adjust shimming stay bosses
Cake discharge is difficult	Inadequate precoat	Prevent feed	Increase precoat feed at 25–40 psig
	Improper conditioning	Conditioner type and dosage	Change conditioner type on dosage based on filter leaf tests
Filter cycle times excessive	Improper conditioning	Chemical dosage	Change chemical dosage
	Feed solids too low	Operation of thickening processes	Improve solids thickening to increase solids concentration in press feed
Filter cake sticks to solids conveying equipment	Change chemical conditioning by using more inorganic chemicals	Conditioning dosage	Decrease ash, increase inorganic conditioners
Procoat pressures gradually increase	Improper sludge conditioning	Conditioning dosages	Change chemical dosage
	Improper precoat feed	Precoat feed	Decrease precoat feed substancially for a few cycles, then optimize
	Filter media plugged	Filter media	Wash filter media
	Calcium buildup in media		Acid wash media (inhibited muriatic acid)
Frequent media blinding	Precoat inadequate	Precoat feed	Increase precoat
	Initial feed rates too high (where no precoat used)		Develop initial cake slowly
Excessive moisture in cake	Improper conditioning	Conditioning dosages	Change chemical dosage
	Filter cycle too short	Correlate filtrate flow rate with cake moisture content	Lengthen filter cycle
Sludge blowing out of press	Obstruction, such as rags, in the press forcing sludge between plates		Shutdown feed pump, hit press closer drive, re-start feed pump-clean feed eyes of plates at end of cycle
Leaks around lower faces of plates	Excessively wet cake soiling the media on lower faces	Cake moisture content	*See* "Excessive moisture in cake"

Source: US EPA.

Table 5
Summary Data—General Information

	Number of plants
Type of biosolids proposed	
Anaerobically digested	21
Anaerobically digested/elutriated	1
Raw	21
Aerobically digested	5
Other (thermally conditioned and chemical)	2
Total	50
Operating status of presses	
In service	41
Abandoned use of press	3
Out of service temporarily	1
Planned to be taken out of service	2
Under startup	2
Under construction	1
Total	50
Filter cake disposal method[a]	
Landfill	24
Incinerate	11
Land apply	5
Incinerate/landfill	3
Landfill/land apply	2
Compost/land apply	1
Total	46
Press manufacturers	
Passavant	14
Edwards & Jones	12
Eimco(Shriver)	11
Sperry	4
Netzsch	3
Hoesch	3
Envirex (NGK)	1
Clow	1
Ingersoil-Rand (Lasta)	1
Total	50
Longest in continuous operation: Statesville, NC–since 1974–Passavant	
Plate material[b]	
Cast Iron	33
Polypropylene	14
Not reported	4
Total	51
Operating experiences[c]	
Positive	29
Negative	12
Mixed	8
Total	49

Source: US EPA.

[a] Excluding plants under construction or which have abandoned use of press.

[b] One plant has both plate-types.

[c] Excluding one plant under construction.

aerobically digested biosolids, one was thermally conditioning anaerobically digested biosolids, and one was dewatering alum biosolids from a tertiary treatment process.

The majority of plants, 41 of them, had their presses in operation at the time of the study. Of the remaining plants, four had taken their presses out of service (one temporarily, three permanently), two planned to take them out of service, two had presses under start-up, and one had not yet completed installation. Most of the plants were landfilling or incinerating their filter cake. Most of the presses used cast iron plates (33 plants), but polypropylene plates were used by some manufacturers for new presses; polypropylene had been used as a replacement for cast iron plates by some plants.

Based on the telephone conversations and on the written comments received from a mail survey, a rating of positive, negative, or mixed was given to the attitude of operations personnel toward operating and maintaining their filter press system. In general, a negative rating was assigned to plants which have taken their press system out of service as a result of excessive operating and/or maintenance costs, which reported serious maintenance problems, or which had a very negative reaction toward the installation. A mixed rating was given to plants which had less serious operating problems and where the operator's attitude was more positive than negative. A positive rating was given where the personnel were generally satisfied or, in some cases, enthusiastic about the press, although some problems may have been reported. The experience of the majority of the plants was rated as positive (29 of 49 plants, excluding one plant under construction). Twelve had negative ratings; of these, four plants have ceased using the press. Eight plants had mixed operating experience.

Table 6 presents a summary of operating comments itemized by type of problem and grouped by the overall rating of operating experience. A large number of plants (28) reported no significant problems with the press installation. It was commonly reported even by plants with both positive and negative overall attitudes toward the press system that costs were high for operating labor and for chemical conditioning, and that well trained, motivated operators and mechanics were a necessity. Plants with positive attitudes seemed to be able to overcome this difficulty by good training and supervision, while plants where the attitude was negative seemed, in contrast, to be overcome by it.

Table 7 presents a summary of data on conditioning chemical dosages and filter cake quality. Most of the plants, 29 of them, were conditioning with ferric chloride and lime only, six were using a precoat of ash or diatomaceous earth with ferric and lime conditioning, and six were conditioning with polymer alone. The less frequent conditioning methods included ferric chloride/lime with ash and polymer/precoat with and without ash. One plant was using lime alone, but was dewatering alum sludge from a tertiary phosphorus removal system.

Reported filter cake solids content averaged 37% for the plants using ferric/lime conditioning and only slightly less at 34% for plants using polymer alone. Plants using a precoat reported one of the highest solids contents, 42% with ferric and lime, but this method had the disadvantage of maximizing performance at the expense of additional inert material in the cake. The two plants using ash as a conditioning material reported 45% with polymer and a precoat and 32% with ferric and lime. However, the amount of ash added, from 63 to 100%, adds significantly to the amount of filter cake to be disposed. Two plants were using polymer with a precoat and reported an average cake

Table 6
Summary of Operating Problems

Item	No. of plants (by overall rating)			Total no. of plants reporting
	Positive	Negative	Mixed	
No significant problem	11	–	–	11
High maintenance costs or unspecified mechanical problems	1	5	2	8
High operating or conditioning costs	4	4	3	11
Well trained and/or motivated operators and mechanics needed	4	4	–	8
Sludge feed problems (line clogs, feed pumps)	2	1	3	6
Presselectrical or instrumentation programs	4	1	–	5
Excessive cloth wear or tears	1	1	4	6
Conditioning system problems (corrosion, line clogs, poor uniformity, or conditioning)	1	2	2	5
Stayboss wear or failure	1	–	3	4
Plate suspension pin breakage	1	–	3	4
Difficult to obtain spare parts	2	2	–	4
Ammonia release problem	1	1	2	4
Plate shifting mechanism problem	2	1	–	3
Cake discharge system problems (conveyors, drip trays)	1	1	1	3
Plate cracking or breakage	–	1	1	2
Hydraulic power unit leakage	–	1	–	1
Poor cake solids	–	1	–	1
Press frame twisting	–	–	1	1
Plate coating wear	1	–	–	1
Rapid cloth blinding	–	1	–	1
Filtrate drain lime accumulation	–	–	1	1

Source: US EPA.

solids of 33%. Table 8 presents a summary of filter press performance on anaerobically digested and raw biosolids for the three most common conditioning methods (ferric/lime, polymer, and ferric/lime/precoat). Surprisingly, in each case, the best results were obtained on digested rather than raw biosolids: 38% vs 36% for ferric/lime, 35% vs 31% for polymer, and 44% vs 32% for ferric/lime/precoat. However, this data should be interpreted carefully, because numerous other factors can influence press performance. Such factors include operating pressure, cloth condition, feed solids percent, primary/secondary biosolids ratio, chemical dosage, and press cycle time (2). An example is Watertown, NY, which reported a 37–44% solids cake with polymer-only conditioning, but required 22.5 kg/T (45 lb/t) of polymer and a very long (4 h) cycle time.

Chemical dosages and cycle times are also listed in Table 8. As shown, chemical dosages for anaerobically digested biosolids averaged (in order by conditioning method): (1) 7% ferric, 26% lime without precoat; (2) 18.5 kg/T (37 lb/t) polymer; and

Table 7
Filter Cake Solids—Average by Conditioning Method

		Solids (%)	
	No. of plants[a]	Mean[b]	Std. dev.
Ferric lime only	29	37	5.3
Ferric/lime/precoat	6	42	7.8
Lime only (alum sludge)	1	38	–
Ferric/lime/ash	1	32	–
Polymer only[c]	6	34	4.2
Polymer/precoat	2	33	4.2
Polymer/ash/precoat	1	45	–

Source: US EPA.
[a] Excluding plants that have abandoned use of press or are under construction.
[b] Using midpoint data for plants reporting a range of values.
[c] Excluding thermally conditioned sludge.

(3) 9% ferric, 32% lime with precoat. Chemical dosages for raw biosolids averaged: (1) 7% ferric, 23% lime without precoat; (2) 6 kg/T (12 lb/t) polymer; and (3) 8% ferric, 20% lime with precoat. Average cycle times varied from about 1.5–2.5 h.

11. GENERAL EQUIPMENT SELECTION CRITERIA

Ohara et al. (36) in writing about the Hyperion system, developed the following set of criteria, in order of priority, for selecting the most cost effective, functional, safe, and environmentally sound system:

a. Meets all environmental and legal requirements.
b. Has minimum energy, resource and economic requirements.
c. Minimum suspended solids remain on the liquid side stream, whether it be concentrate, filtrate, or supernatant.
d. Provides capture of solids.
e. Provides maximum cake solids (minimum percent moisture in the cake).
f. Has maximum operational reliability, flexibility, and ease of use.
g. Requires minimum maintenance and downtime.
h. Has maximum flexibility to meet changing needs.
i. Can meet established construction schedules.

12. COSTS

Filter presses carry relatively high capital costs compared with other mechanical dewatering methods because of equipment and the need for standby capability for cake handling. Operation and maintenance (O&M) costs might also be relatively high. O&M cost elements include chemicals (for biosolids conditioning and precoating), cloth washing and replacement, and operator activity (1,37).

Figure 7 gives fixed-volume, recessed plate pressure filter capital cost as a function of press volume. Costs include those for filter auxiliary equipment, piping, and building (26). All costs are in 1975 USD (Cost Index = 190.49). To obtain the values in terms of the present 2006 USD using the Cost Index for Utilities shown in Appendix, multiply the costs by a factor of 2.76 (38). As an example, a pressure filter having 100 ft^3 (2.8 m^3)

Table 8
Filter Cake Solids–Average by Conditioning Methods and Biosolids Type

Conditioning method and biosolids type	No. of plants reporting	Average[a] cake solids (%)	Ferric only (%)			Lime only (%)			Polymer only (lb/t)			Cycle time[b] (min)		
			Min.	Max.	Avg.	Min.	Max.	Avg.	Min.	Max.	Avg.	Min.	Max.	Avg.
Ferric/lime														
Anaerobic	11	38	3	15	7	12	43	26				52	168	100
Raw	14	36	1.5	12	7	15	30	23				70	330	155
Polymer														
Anaerobic	2	35							20	45	37	60	240	150[c]
Raw	3	31							7	18	12	50	105	82
Ferric/lime/precoat														
Anaerobic	4	44	7	10	9	16	40	32				90	105	99
Raw	1	32	–	–	8	–	–	20				–	–	120

Source: US EPA.

[a] Using midpoint of data for plants reporting a range of values.

[b] Excuding diaphragm press installations.

[c] Only two plants reporting, wide variation.

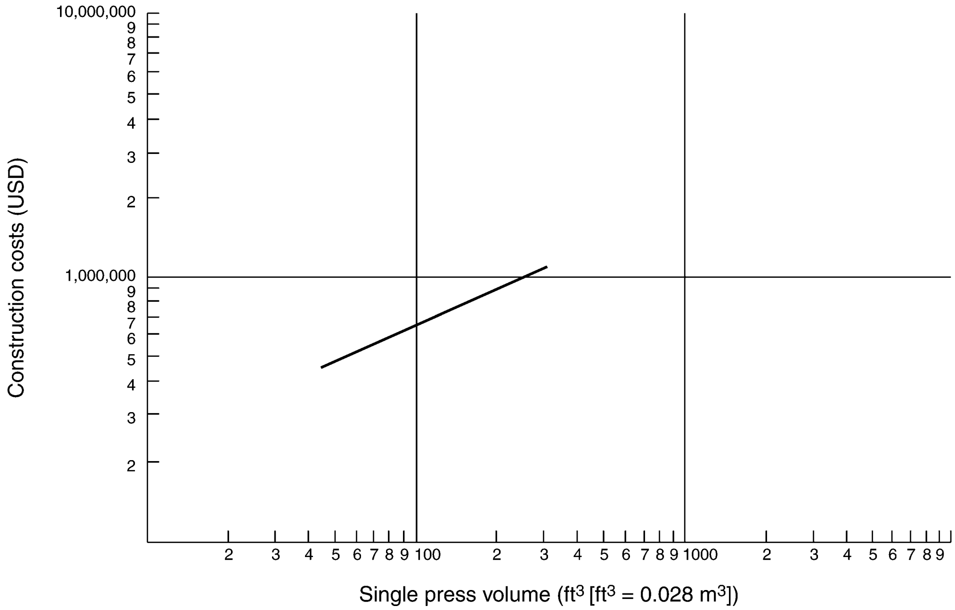

Fig. 7. Costs for fixed-volume recesses plate pressure filter press (1975) (*Source:* US EPA).

capacity would cost about 700,000 USD. Since this number is based on 1975 cost, it must be adjusted to the current design year, which gives a cost equal to 2006 USD 1,935,000.

Figure 8 indicates fixed-volume, recessed plate pressure filter labor requirements (26). Labor requirements are based on continuous, 7 d/wk operation with 2-h cycles and include operation and maintenance for both press and related auxiliaries (chemical feed system and pumps). As an example, a pressure filter having 100 ft^3 (2.8 m^3) of capacity would require 8000 man-hour of operation and maintenance per year and would be included in the cost analysis.

Figure 9 gives power consumption as a function of feed solids concentration and operating volume (26). The graph is based on a filter that operates continuously, 7 d/wk, and has a 2-h cycle time. Power consumption is included for the feed pump, open and close mechanisms, and moveable head mechanism.

Figure 10 presents a graph developed for estimating annual material and maintenance costs for a fixed-volume, recessed plate pressure filter. The graph is based on unit operation of 7 d/wk with a 2-h cycle time (26). All costs are in 1975 USD (Cost Index = 190.49). To obtain the values in terms of the present 2006 USD using the Cost Index for Utilities shown in Appendix, multiply the costs by a factor of 2.76 (38).

13. DESIGN EXAMPLES

13.1. Design Example 1

This is a design example with supporting calculations for sizing a fixed-volume recessed plate filter press for the design of the biosolids dewatering system. Use the following assumptions and criteria (10):

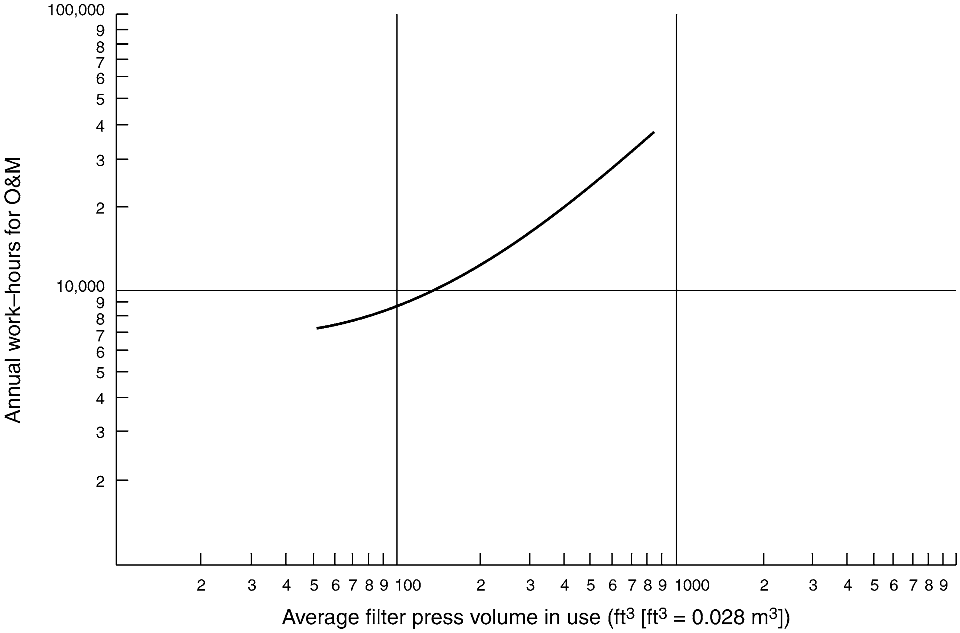

Fig. 8. Annual OM man-hour requirements for fixed-volume recessed plate pressure filter press (*Source:* US EPA).

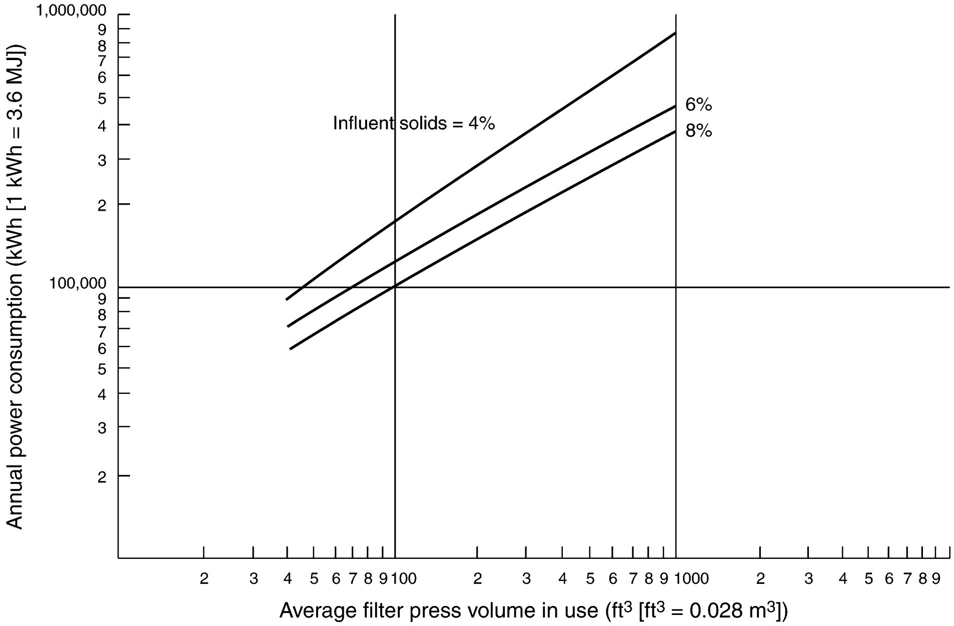

Fig. 9. Power consumption for fixed-volume recessed plate pressure filter press (*Source:* US EPA).

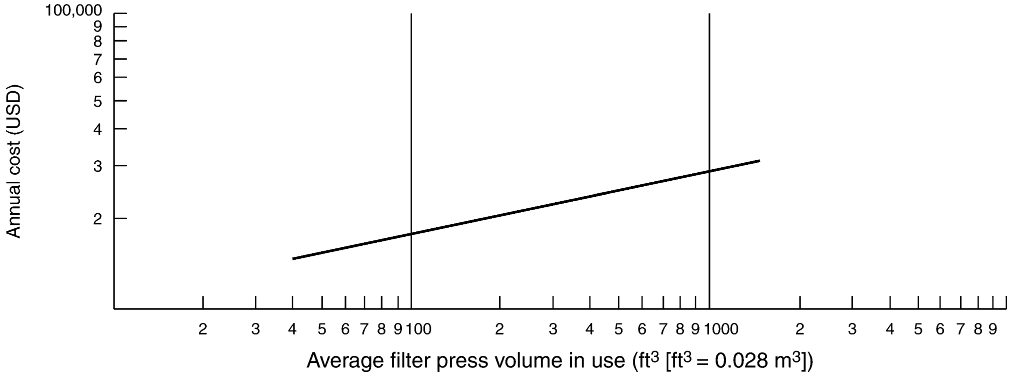

Fig. 10. Annual maintenance material cost for fixed-volume recessed plate pressure filter press (1975) (*Source:* US EPA).

The characteristics of the biosolids stream are given as follows:

a. Type of solids: chemical/biological.
b. Design biosolids flow = 37,850 L/d (10,000 gpd).
c. Concentration of solids = 2%.
d. Specific gravity of feed = 1.

The minimum dry solids allowed in cake will be 30% by weight.
The following treatability testing design data is given:

1. Biosolids cake characteristics.
   a. Cake thickness = 32 mm (1.25 in.).
   b. Wet cake density = 1280 kg/m^3 (80 lb/ft^3).
2. Optimum chemical conditioning.
   a. Lime dosage (CaO) = 10% by weight of dried solids.
   b. Ferric chloride dosage ($FeCl_3$) = 5% by weight of dried solids.
3. Dewatering equipment requirements.
   a. Operating time = 8 h/d, 5 d/wk.
   b. Cycle time (variable flow/pressure system).
      i. Feed = 30 min (1800 s) at 172 kPa (25 psig).
      ii. Feed = 30 min (1800 s) at 345 kPa (50 psig).
      iii. Feed = 30 min (1800 s) at 517 kPa (75 psig).
      v. Feed = 1 min (60 s) at 690 kPa (100 psig).
      v. Cake discharge = 29 min (1740 s).
      vi. Total = 120 min (7200 s) or 4 cycles/d.

The number of filter press units will be selected such that 100% of the design liquid biosolids flow rate is filtered when the largest single unit is out of service. This example will also assume that the maximum filtration capacity shall be 125% of the design capacity when all units are in operation. Any additional specific optional features or supporting systems (i.e., precoating, air blowing, and filter media wash systems) and associated sizing requirements will be determined once the filter press has been selected.

a. Determine required filter volume.
b. Select efficient filter units.

c. Compute the size of biosolids storage required.
d. Compute the size of lime storage facility required.
e. Compute the size of the ferric chloride storage facility required.
f. Compute the size of the conditioning tank.

Solution

1. Determine required filter volume.

   Daily dry solids generation rate = [(37,850 L/d) × (0.02) × (1) × (1 kg/L)] = 760 kg/d (1670 lb/d).
   Total dry solids dewatered = biosolids + lime + ferric chloride.
   Solids = (760 kg/d × 7 d/wk)/(5 d/wk) operation = 1064 kg/d.
   Lime (CaO) = 1064 kg/d × (0.10) = 106 kg/d.
   Ferric chloride = 1064 kg/d × (0.05) = 53 kg/d.
   Total dry solids/d = 1046 + 106 + 53 = 1223 kg/d ~ 1230 kg/d (2700 lb/d).
   Filter biosolids volume required per cycle = [1230 kg/d]/[(4 cycle/d) × (1280 kg/m^3) × (0.30)] = 0.8 m^3 (29 ft^3).

2. Select efficient filter units.

   Determine the pressure filter sizes available. From the manufacturers' catalogs (39), determine the sizes of various filter units. Tabulate the filter area available with and without the single largest unit (*see* Table 9). Select proper filter press unit. The most efficient and manageable filter press unit assembly is the one that has the fewest operating units and provides nearly 100% operating capacity when one unit is out of service, and about 25% extra capacity when all units are in operation. Based on this method, the proper unit selection from Table 9 would be Unit F. This selection has a total of four units, including three operating units and one standby unit. This assembly will provide 105% of the design is daily requirement when one unit is not operating and 140% of the design is daily dewatering capacity when the standby unit is in operation. Although Unit D required the same number of units and has the same overall operating capacities, this selection is the maximum capacity of this size unit and if necessary, would not allow for additional future capacity.

   Although the method presented provides a direct approach to selecting the optimal size and number of required presses, an economic and technical evaluation of several alternatives which achieve the minimum requirements should be considered before the final selection of the appropriate filter size and number of associated units.

3. Compute the size of biosolids storage required.

   Assume biosolids must be stored for 4 d.
   Volume of biosolids to be stored = 37,850 *L*/d × 4 d = 151,400 *L* (40,000 gal).
   Therefore, a storage tank or tanks should be selected with a capacity of 152,000 *L* or 152 m^3 (40,000 gal) from manufacturers' catalogs.

4. Compute the size of lime storage facility required.

   Provide 30 d or 1 mo of storage. Assume approx 4.33 wk/mo.
   Lime required per 8-h/d = 106 kg/d.
   Lime required per month = 106 kg/d × 5 d/wk × 4.33 wk/mo = 2300 kg/mo (5070 lb/mo).
   Assume use of hydrated lime at 96% purity.
   For 30 d storage quantity of hydrated lime required.

Table 9
Selection Chart of Recessed Fixed-Volume Plate and Frame Filter Press Units

Unit I.D.	Filter size (mm)[a]	No. of chambers	Volume of each unit (m^3)	Minimum units required	Total volume with minimum units (m^3)	Total units with one standby	Total volume with one standby unit (m^3)	Volume without standby unit (%)[b]	Volume with standby unit (%)[c]
A	470	27	0.11	8	0.88	9	0.99	110	124
B	470 (max.)	40	0.20	5	1	6	1.2	125	150
C	630	27	0.23	4	0.92	5	1.15	115	144
D	630 (max.)	30	0.28	3	0.84	4	1.12	105	140
E	800	16	0.23	4	0.92	5	1.15	115	144
F	800	20	0.28	3	0.84	4	1.12	105	140
G	800 (max.)	40	0.57	2	1.14	3	1.71	143	214

Source: US EPA.

Conversions: 1 $m^3 = 35.3$ ft^3, 1 in. = 25 mm.

[a] Data compiled from actual manufacturer's data.

[b] Total volume without standby unit/actual volume required.

[c] Total volume with one standby unit/actual volume required.

= [(2300 kg/mo)/0.96] × [74/56].

= 3170 kg/mo (7160 lb/mo).

Number of bags of hydrated lime needed per month = (3170 kg/mo)/(45 kg/bag) = 71 bags/mo.

Therefore, storage should be provided for 80 bags.

The capacity of the hydrated lime feed hopper should be large enough to contain 1 d supply of lime.

Bags needed per day = [71 bags/mo]/[(5 d/wk) × (4.33 wk/mo)] = 4 bags/d

5. Compute the size of the ferric chloride storage facility required.
   Provide 30 d or 1 mo of storage at operating condition.
   Ferric chloride required per 8-h/d = 53 kg/d.
   Ferric chloride required per month = 53 kg/d × 5 d/wk × 4.33 wk/mo = 1150 kg/mo (2530 lb/mo).
   Assume use of ferric chloride at 40% purity with a density of 1.45 kg/L (12 lb/gal).
   For 30-d storage, quantity of ferric chloride required = [(1150 kg/mo)/(0.40 × 1.45 kg/L] = 1990 L/mo (530 gal/mo).
   Therefore, minimum storage should be provided for 2000 L or 2 m^3 (550 gal).
6. Compute the size of the conditioning tank.
   Assume use of an in-line conditioning tank with a 10 min (600 s) detention and mixing time.
   Volume required for conditioning tank = 2.2 L/s × 600 s = 1320 L (350 gal).
   Therefore, a conditioning tank should be selected with a minimum capacity of 1320 L or 1.32 m^3 (350 gal). This conditioning tank should also be equipped with a mixer and level switches to control the operation of the biosolids and dilute conditioning chemical metering pumps.

13.2. Design Example 2

This problem provides another design example and supporting calculations for sizing a variable-volume recessed plate filter press (40) for a biosolids dewatering system. The following assumptions and criteria are to be used (10).

Characteristics of the biosolids stream.

Type of solids: biological.
Designed daily liquid biosolids flow = 80,000 L/d (21,100 gpd).
Concentration of solids = 5%.
Specific gravity of feed = 1.

The minimum dry solids allowed in the cake will be 25% by weight.

Treatability testing has been done and the following design data was obtained:

1. Sludge cake characteristics.
   a. Cake thickness = 32 mm (1.25 in.)
   b. Wet cake density = 1120 kg/m^3 (70 lb/ft^3).
2. Optimum chemical conditioning.
   a. Lime dosage (CaO) = 5% of dried solids by weight.
   b. Organic polymer = 2% of dried solids by weight.
3. Dewatering equipment requirements.
   a. Operating time = 8 h/d, 5 d/wk.
   b. Cycle time.
      i. Feed = 20 min (1200 s) at 345 kPa (50 psig).
      ii. Compression = 15 min (900 s) at 1550 kPa (225 psig).

iii. Cake discharge = 25 min (1500 s).
iv. Total = 60 min (3600 s) or 8 cycles/d.

The number of filter press units will be selected such that 100% of the designed liquid biosolids flow rate is filtered when the largest single unit is out of service. This example will also assume that the maximum filtration capacity shall be 125% of the design capacity when all units are in operation. Any additional specific optional features or supporting systems (i.e., precoating, air blowing, and filter media wash systems) and associated sizing requirements will be determined once the filter press has been selected.

a. Determine required filter volume.
b. Select efficient filter press units.
c. Compute the size of biosolids storage required.
d. Compute the size of lime storage facility required.
e. Compute the size of the polymer storage facility required.
f. Compute the size of the conditioning tank.

Solution

a. Determine required filter volume.
Total daily dry solids generation rate = [(80,000 L/d) × (0.05) × (1) × (1 kg/L)] = 4000 kg/d (8800 lb/d).
Total dry solids dewatered per day = biosolids + lime + polymer.
Sludge solids = [4000 kg/d × 7 d/wk]/[5 d/wk (operation)] = 5600 kg/d.
Lime (CaO) = 5600 kg/d × (0.05) = 280 kg/d.
Polymer = 5600 kg/d × (0.02) = 112 kg/d.
Total dry solids per day = 5600 + 280 + 112 = 5992 kg/d ~ 6000 kg/d (13,200 lb/d).
Filter biosolids volume per cycle = [6000 kg/d]/[(8 cycle/d) × (1120 kg/m^3) × (0.25)]
= 2.7 m^3 (95 ft^3).

b. Select efficient filter press units.

Determine the pressure filter sizes available: Examine the manufacturers' catalogs (39) to determine the sizes of various filter units. Tabulate the filter area available with and without the single largest unit (*see* Table 10).
Select proper filter press unit: The most efficient and manageable filter unit assembly is the one that has the fewest operating units and provides nearly 100% operating capacity when one unit is out of service, and about 25% extra capacity when all units are in operation. Based on this method, the proper unit selected from Table 10 would be Item D. This selection has a total of five units, including four operating units and one standby unit. This assembly will provide 104% of the designed daily requirement when one unit is not operating and about 133% of the designed daily dewatering capacity when the standby unit is in operation. Although Item C has the same number of units and has the same overall operating capacities, this selection is the maximum capacity of this size unit and if necessary, would not allow for additional future capacity.
Although the method presented provides a direct approach for selecting the optimal size and number of required presses, an economic and technical evaluation of several alternatives that achieve the minimum requirements should be considered prior to the final selection. For example, by comparing different alternatives in Table 10, Item C and Item D are both similar for the number of units and operating capacities. However, although Item C is at its maximum size, it has some additional capacity (i.e., 105%) and would be less expensive than Item D and may be suitable if no additional capacity is required.

Table 10
Selection Chart of Recessed Variable-Volume (Diaphragm) Plate and Frame Filter Press Units

Unit I.D.	Filter size (mm)[a]	No. of chambers	Volume of each unit (ft^3)	Minimum units required	Total volume with minimum units (ft^3)	Total units with one standby	Total units with one standby unit (ft^3)	Volume without standby unit (%)[b]	Volume with standby unit (%)[c]
A	630 (max.)	33	0.28	10	2.8	11	3.1	104	115
B	800 (min.)	16	0.23	12	2.8	13	3	104	111
C	800 (max.)	53	0.71	4	2.8	5	3.6	104	133
D	1000 (min.)	31	0.71	4	2.8	5	3.6	104	133
E	1000 (max.)	62	1.42	2	2.8	3	4.3	104	159
F	1200 (min.)	62	2.12	2	4.2	3	6.4	156	237

Source: US EPA.

Conversions: 1 m^3 = 35.3 ft^3, 1 in. = 25.4 mm.

[a] Data compiled from actual manufacturer's data.

[b] Total volume without standby unit/actual volume required.

[c] Total volume with one standby unit/actual volume required.

c. Compute the size of biosolids storage required.
   Assume that the biosolids must be stored for 4 d.
   Volume of biosolids to be stored = 80,000 L/d × 4 d = 320,000 L (84,400 gal).
   Therefore, a storage tank or tanks should be selected with a capacity of 320,000 L or 320 m^3 (85,000 gal) from manufacturers' catalogs.

d. Compute the size of lime storage facility required.
   Provide 30 d or 1 mo storage and assume 4.33 wk/mo and the use of hydrated lime at 90% purity.
   Lime required per 8 h/d = 280 kg/d (620 lb/d).
   Lime required per month = 280 kg/d × 5 d/wk × 4.33 wk/mo = 6100 kg/mo (13,430 lb/mo).
   For 30 d storage quantity of hydrated lime required = [(6100 kg/mo)/0.9] × (74/56) = 896 kg/mo (1974 lb/mo).
   Number of bags of hydrated lime needed per month = (8960 kg/mo)/(45 kg/bag) = 199 bags/mo. Therefore, storage should be provided for 200 bags. The capacity of the hydrated lime feed hopper should be large enough to contain 1 d supply of lime.
   Number of bags needed per day = (199 bags/mo)/[(5 d/wk) × (4.33 wk/mo)] = 10 bags/d.
   Based on the bulk density of hydrated lime of 480 kg/m^3 (30 lb/ft^3) and allowing provision for two hoppers to hold 5 bags of hydrated lime each.
   Volume required for each hopper = [(5 bags) × (45 kg/bag)]/(480 kg/m^3) = 0.47 m^3 (17 ft^3).
   Therefore, each hopper should have a minimum capacity of 0.5 m^3 (17 ft^3) and will feed lime into a mixing tank where, a 10% lime slurry by weight will be prepared before being metered into the biosolids conditioning tank.

e. Compute the size of the polymer storage facility required.
   Provide 30 d or 1 mo storage.
   Polymer required per 8-h d = 112 kg/d (250 lb/d).
   Polymer required per month = 112 kg/d × 5 d/wk × 4.33 wk/mo = 2425 kg/mo (5420 lb/mo).
   Number of bags of polymer needed per month = (2425 kg/mo)/(22 kg/bag) = 110 bags/mo.
   Therefore, storage should be provided for 110 bags.
   The capacity of the polymer feed hopper should be large enough to contain 1 d supply of polymer.
   Number of bags needed per day = (112 kg/d)/(22 kg/bag) = 5 bags.
   The bulk density of the polymer is assumed to be 320 kg/m^3 (20 lb/ft^3) and the daily supply will be stored in two hoppers.
   Volume required for each hopper = (112 kg/d)/(2 × 320 kg/m^3) = 0.2 m^3 (7 ft^3)
   Therefore, each hopper should have a minimum capacity of 0.2 m^3 (7 ft^3) and will feed polymer to a mixing tank for preparation of the 5% solution before being metered into the biosolids conditioning tank.

f. Compute the size of the conditioning tank.
   Assume use of an in-line conditioning tank with a 10-min (600-s) retention and mixing time.
   Volume required for conditioning tank = 12 L/s × 600 s = 7200 L (1900 gal)

   Therefore, a conditioning tank should be selected with a 7200 L or 7.2 m^3 (1900 gal) working capacity.
   This conditioning tank should also be equipped with a mixer and level switches to control the operation of the biosolids and conditioning chemical metering pumps.

NOMENCLATURE

a and *b*	Coefficients
A	Area of cake
A	Filtration area

c	Weight of dry solids per unit volume in the unfiltered slurry
c	Feed solids concentration
c	Mass of cake deposited per unit volume of filtrate
d	Diameter of capillary
$dv/d\theta$	Rate of flow of liquid across cake
g	Gravitational constant
K	A constant equal to 5
ℓ	Cake thickness (cm)
L	Length of capillary
L	Cake thickness
P	Pressure differential
P_c	Pressure drop across cake
P_t	Total pressure drop across cake and medium
r	Specific resistance of cake
R	Specific resistance (m/kg)
R	Specific resistance (s^2/g)
R_m	Resistance of the medium
S_o	Specific surface
S_c	Cake solids concentration by weight fraction
T	Tonne (Metric)
t	Ton (English)
t	Time
U	Linear velocity
v	Volume of solids deposited per unit of filtrate
V	Volume of filtrate
ε	Porosity
φ	Filtrate flow rate or flux
μ	Absolute viscosity
ρ_t	Feed density (g/cm^3)
θ	Filtration time (min)

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APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers[a]

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

[a] From ref. 38.

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Evaporation Processes

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CONTENTS

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1. INTRODUCTION

1.1. *Drying and Evaporation Processes*

Water removal from municipal and industrial effluent streams constitutes an important step in wastewater and sludge treatment. The purpose is to concentrate, separate, dispose, or utilize wastes and pollutants and to regenerate and return clean water to the environment. In this context, the discussion here will be limited only to industrial and municipal sludge dewatering, evaporation, and drying (1–5).

Although sand bed dewatering of sludge has been popular in small communities, heat drying or evaporation have proved feasible in many instances. Water evaporation and heat drying are currently expensive and require fuel consumption to remove the water. They become feasible when the dried sludge can be sold as a fertilizer or used as a vitamin- and protein-enriched animal feedstock. Such possibilities depend not only on the regular market, but also on the attitude of the public to demand recycling of wastes and accept the associated cost. Sludge return to the environment in a dry form and utilization of its nutrient content may be ecologically more attractive than the current trend toward incineration. Although the latter uses the heat content of the sludge to accomplish combustion, the problem of ash disposal remains.

Recent incineration practice usually involves prior removal of excess water and sludge thickening through dewatering and drying steps. Preliminary drying of the sludge may take place either in a separate unit or in the first section of the incinerator. Furthermore, incineration equipment is often designed so as to permit flexibility of operation during either drying or combustion. The significance of water removal steps

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in various sludge and wastewater treatment designs, the associated energy demands and costs, and the possible improvement of the market for dried sludge warrant analysis and further evaluation of drying and evaporation.

A rigid, sharp distinction among evaporation, dewatering, and drying does not exist. All involve water removal to some extent, and also reduction of the weight and volume of the fluid sludge to that of a concentrated sludge or a moist solid. Usually the water content of a dewatered sludge is higher than that of dried sludge. However, sand beds used for physical dewatering are called "sludge drying beds," despite the considerable water content of the remaining cake, in order to distinguish them from other mechanical dewatering systems. In general, water is removed in drying beds and lagoons by natural drainage and evaporation. Drying to low water content requires high-temperature water removal in mechanical dryers. Depending on the drying temperature, simultaneous sludge sterilization can be achieved.

The origin of the effluent stream, the final desirable moisture content, and the end use of the sludge determine whether dewatering is sufficient, or heat drying is necessary. Because, thermal drying is usually more expensive than physical or mechanical water removal, dewatering prior to heat drying is desirable.

Before attempting any theoretical analysis of drying and evaporation processes, the design and operating characteristics of some basic units will be presented. This will familiarize the reader with the operation of physical and mechanical dryers and evaporators and will permit a rational theoretical treatment of these systems. Drying and evaporation consist of a combination of mass and heat transfer processes common to all dryer designs for a given sludge.

1.2. Natural Sludge Evaporation Lagoons and Evaporation Process Reactor

In this chapter, both natural and man-made evaporation processes will be discussed. Evaporation can be defined as the process by which liquid water is converted into a gaseous state. Evaporation can only occur when water is available. It also requires that the humidity of the atmosphere be less than the evaporating surface (at 100% relative humidity there is no more evaporation). The evaporation process requires large amounts of energy. For example, the evaporation of 1 g of water at a temperature of 100°C requires 540 calories of heat energy (600 calories at 0°C) (43).

Because the design and operation of natural sludge evaporation lagoons is presently more of an art than a science, a number of empirical design criteria and variables will be discussed here rather than deferring them to theoretical discussions. Comparison of drying periods, land requirements, and solids loading rates between sand beds and lagoons obviously favor the dewatering technique. However, lagoons are quite commonly used for sludge drying where inexpensive land is available, because of their simple, low-cost operation. Precipitation and evaporation rates are the controlling factors (6,7).

Several large cities have used lagoon drying successfully for several years. Most organic industrial sludges are often dried in lagoons since offensive odors are minimal. Lagooning has been proved to be economical for oil and metal finishing sludges for which vacuum filtration is difficult.

The theory, principles, and design of thermal evaporation process reactors are well established and they are presented in Section 3.

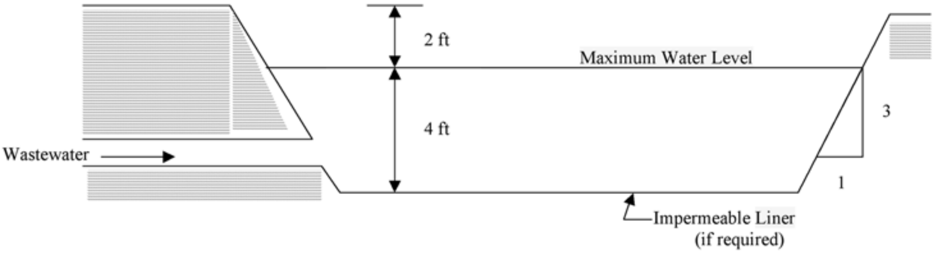

Fig. 1. The sludge evaporation lagoon.

2. SLUDGE EVAPORATION LAGOONS (SLUDGE DRYING LAGOONS)

2.1. Process Description

2.1.1. General Process Description

The sludge evaporation lagoon (Fig. 1) may be described as an open holding facility which depends solely on eliminating conditions such as evaporation, precipitation, temperature, humidity, and wind velocity to effect dissipation (evaporation) of on-site wastewater. Individual lagoons may be considered an alternate means of wastewater disposal on individual pieces of property. The basic impetus to consider this system is to allow building and other land uses on properties, which have soil conditions not conducive to the workability and acceptability of the traditional on-site drainfield or leachbed disposal systems (1–3).

Generally if the annual evaporation rate exceeds the annual precipitation, this method of disposal may at least be considered. The deciding factor then becomes the required land area and holding volume. It should be noted that for unlined on-site installation such as homes and small industrial applications, there might also be a certain amount of infiltration or percolation in the initial period of operation. However, after a time, it may be expected that solids deposition will eventually clog the soil surface to the point where infiltration is eliminated. The potential impact of wastewater infiltration to the groundwater, and particularly on-site water supplies, should be evaluated in any event and, if necessary, lagoon lining may be utilized to alleviate the problem.

Lagoons are often preceded by septic tanks or aerobic units in order to provide a more acceptable influent to and minimize sludge removal from the lagoon (5).

2.1.2. Process Operation

Sludge is placed in the lagoon at depths three to four times greater than it would be in a drying bed. Generally, sludge is allowed to dewater and dry to some predetermined solids concentration before removal, a process that might require 1–3 yr. The cycle is then repeated. Sludge should be stabilized prior to addition to the lagoon to minimize odor problems. Large areas of lagoons can produce nuisance odors as they go through a series of wet and dry conditions.

Sludge drying lagoons consist of retaining walls, which are normally earthen dikes 2–4 ft (0.7–1.4 m) high. The earthen dikes normally enclose a rectangular space with a permeable surface. Ancillary equipment includes sludge feed lines and metering pumps, supernatant decant lines, and some type of mechanical sludge removal equipment. The removal equipment may include a bulldozer, dragline, or front-end loader.

In areas where permeable soils are unavailable, underdrains and associated piping may be required.

Operating procedures common to all types of sludge drying lagoons involve:

(a) Pumping liquid sludge, over a period of several months or more, into the lagoon. The pumped sludge is normally stabilized prior to application. The sludge is usually applied until a lagoon depth of 24–48 in. (0.7–1.4 m) is achieved.
(b) Decanting supernatant, either continuously or intermittently, from the lagoon surface and returning it to the wastewater treatment plant.
(c) Filling the lagoon to a desired sludge depth and then permitting it to dewater. Depending on the climate and the depth of applied sludge, the time involved for dewatering to a final solids content of between 20% and 40% solids may be 3–12 mo.
(d) Removing the dewatered sludge with some type of mechanical removal equipment.
(e) Resting (adding no new sludge) the lagoon for three to six months.
(f) Repeating the cycle.

2.2. Process Applications and Limitations

2.2.1. Applications

The "technology" of evaporation is well developed in terms of scientific understanding and application of climatological and meteorologic data.

The on-site utilization of evaporation lagoons for the disposal of domestic wastewater, from homes and smaller industrial or commercial facilities, may be applicable where access to a municipal sanitary sewer is not available, where subsurface methods are not feasible, and where effluent polishing for surface discharge is not practical.

2.2.2. Limitations

The limitations for evaporation lagoons include local health ordinances; potential for odors and health hazard when not properly designed; land area requirements; dependence on meteorological and climatological conditions. Sludge lagoons may require provisions to add makeup water to maintain a minimum depth during dry, hot seasons. Finally, public access restrictions are necessary.

2.2.3. Advantages and Disadvantages

The advantages of using sludge evaporation lagoons (or sludge drying lagoons) are

(a) Lagoons are low-energy consumers.
(b) Lagoons consume no chemicals.
(c) Lagoons are not sensitive to sludge variability.
(d) The lagoons can serve as a buffer in the sludge handling flow stream. Shock loadings due to treatment plant upsets can be discharged to the lagoons with minimal impact.
(e) Organic matter is further stabilized.
(f) Of all the dewatering systems available, lagoons require the least amount of operation attention and skill.
(g) If land is available, lagoons have a very low capital cost.

The disadvantages of using sludge evaporation lagoons are

(a) Lagoons may be a source of periodic odor problems, which may be difficult to control.
(b) There is potential for pollution of groundwater or nearby surface water.
(c) Lagoons can create vector problems (for example, flies and mosquitoes).

(d) Lagoons are more visible to the general public.
(e) Lagoons are more land-intensive than fully mechanical methods.
(f) Rational engineering design data are lacking to allow sound engineering economic analysis.

2.3. Design Considerations

2.3.1. Target Process Performance

The performance of evaporation lagoons is necessarily site-specific; therefore, the following data are presented on the basis of net annual evaporation rate that may exist in a certain area:

Net annual evaporation evaporation–precipitation (in.)	Lagoon performance water evaporation (gal/ft^2/yr)
5	3.1
10	6.2
15	9.4
20	12.5
40	24.9
60	37.4

Periodic pump out of accumulated sludge is required from pretreatment unit and/or lagoon.

Lagoon dewatering of sludges does not usually result in fork-liftable sludge. Dewatering from 5% solids to 40–45% solids lasts 2–3 yr, and a 3-yr cycle is usually recommended for lagoon dewatering. Sludge is first dewatered in a lagoon for 1 yr. The lagoon is then allowed to dry for 12–18 mo, followed by a rest period of 6–12 mo.

2.3.2. Design Criteria

Proper design of sludge drying lagoons requires a consideration of the following factors: climate, subsoil permeability, sludge characteristics, lagoon depth, and area management practices. A detailed discussion of these factors follows.

2.3.2.1. CLIMATE

After dewatering by drainage and supernating, drying in a sludge lagoon depends primarily on evaporation. Proper lagoon design, therefore, requires climatic information concerning:

(a) Precipitation rate (annual and seasonal distribution).
(b) Evaporation rate (annual average, range, and seasonal fluctuations).
(c) Temperature extremes.

2.3.2.2. SUBSOIL PERMEABILITY

The subsoil should have a moderate permeability of 1.6×10^{-4} to 5.5×10^{-4} in./s (4.2×10^{-4} to 1.4×10^{-3} cm/s) and the bottom of the lagoon should be a minimum of 18 in. (46 cm) above the maximum groundwater table, unless otherwise directed by local authorities.

2.3.2.3. Sludge Characteristics

The type of sludge to be placed in the lagoon can significantly affect the amount and type of odor and vector problems that may be produced. It is recommended that only those sludges that have been anaerobically digested be used in drying lagoons.

2.3.2.4. Lagoon Depth and Area

The actual depth and area requirements for sludge drying lagoons depend on several factors, such as precipitation, evaporation, type of sludge, volume, and solids concentration. Solids loading criteria have been given as 2.2–2.4 lb of solids/yr/ft^3 (36–39 kg/yr/m^3) of capacity. A minimum of two separate lagoons, or even three lagoons, is provided to ensure availability of storage space during cleaning, maintenance, or emergency conditions.

2.3.2.5. Structure

Lagoons may be of any shape, but a rectangular shape facilitates rapid sludge removal. Lagoon dikes should have a slope of 1:3, vertical to horizontal, and should be of a shape and size to facilitate maintenance, mowing, passage of maintenance vehicles atop the dike, and access for the entry of trucks and front-end loaders into the lagoon. Surrounding areas should be graded to prevent surface water from entering the lagoon. Return must exist for removing the surface liquid and piping to the treatment plant. Provisions must also be made for limiting public access to the sludge lagoons.

2.3.2.6. Hydraulic Loading

The hydraulic loading is the primary sizing criteria for an individual home total-retention lagoon. In order to size the system properly, the following information is needed:

(a) Anticipated flow of wastewater.
(b) Evaporation rates (10-yr minimum of monthly data).
(c) Precipitation rates (10-yr minimum of monthly data).

The rate of wastewater flow is expected to be in the range of 50 gal per capita per day, depending on the individual site location. Precipitation and evaporation data for most areas can be readily found in weather records. A 12-mo mass balance should be utilized to properly determine design sizing. Design criteria include a depth of 2–4 ft, level bottoms, and banks more than 2 ft higher than maximum water level.

2.3.2.7. Monitoring Design

Table 1 summarizes the process variables, measurements, and the monitoring instruments recommended by the US Environmental Protection Agency (US EPA).

Operation of sludge evaporation lagoons is generally slow, proceeding mainly through evaporation. Because of the longer retention of water in lagoons, sludge stabilization is necessary to minimize noxious odors.

Factors discussed in relation to sand bed design, such as climatic conditions, sludge properties, subsoil permeability, sludge load, and so on, also determine the design of lagoons. The design should provide for at least two, or even three, lagoons having a maximum depth of 4 ft. Depending on the climate and the sludge characteristics, 1–4 ft^2/capita are required for sludge drying. A solid loading rate of about 2.4 lb/ft^3/yr (39 kg/m^3/yr) is recommended.

Table 1
Process Variables, Measurements, and Instruments of Sludge Evaporation Lagoons

Process variables	Measurements	Instruments
Feed sludge	Flow	Venturi with diapragm sensors Magnetic Doppler Pump displacement
	Pressure	Bourdon with cylindrical seal
	Density	Nuclear Optical Ultrasonic
Lagoon contents	Moisture content	Portable ohmmeter Lab test
Harvested sludge	Flow (volume)	Transport displacement
	Weight	Static
Supernatant and surface runoff		
Weather	Wind speed [15 ft (4.6 m)] above ground	Anemometer
	Wind direction [15 ft (4.6 m)] above ground	Wind vane
	Temperature [5 and 25 ft (1.5 and 7.6 m)] above ground	
	Relative humidity	RTD with solar shield Thermistor with solar shield
	Rainfall	RTD with lithium chloride cloth (wet bulb temperature)
	Solar radiation	Tipping bucket Thermophile
Atmospheric monitoring	Odors	Portable olfactometer

Source: US EPA.

2.3.3. *Environmental Impact and Energy Consumption*

Process reliability is good. However, it should be closely controlled to prevent a health hazard. Potential odors, health hazards, and large land area requirements may adversely affect surrounding property value.

Lagoons are usually gravity fed from the source. Where pumping is required and assuming a wire to water efficiency of 60%, using the following expression would approximate energy requirements:

$$\text{kwh/yr} = 0.0019\ (\text{flow in gal/d})\ (\text{discharge head in ft})$$

2.4. *Cost*

2.4.1. *Construction Cost*

The US EPA published information on capital cost of constructing sludge lagoons. Using an ENR CC Index equal to 6390.21 (January 2002 Cost), typical excavation and liner costs associated with a two-bedroom residence are as follows:

Item	Unit price	Cost US $
Excavation and hauling (750 yd^3)	\$1.96/yd^3	1472
Liner, 10 mil PVC (21,000 ft^2)	\$0.28/ft^2	5964
Supervision and hand labor	—	2117
Total		9553

To the above must be added costs for land, fencing, septic tank, and ancillary items.

Table 2
Sludge Drying Lagoons, Labor Requirements

Dry solids applied (tons/yr)	Labor (h/yr)		
	Operation	Maintenance	Total
100	30	55	85
1000	55	90	145
10,000	120	300	420
50,000	450	1500	1950

2.4.2. *Operating Costs*

Septic tank pump out is the only energy cost. Pumping of a septic tank is estimated to be about US \$25.82/yr.

Table 2 indicates labor requirements for sludge drying lagoons. The requirements include periodic removal of solids and minor maintenance requirements, such as dike repair and weed control. No information is available on maintenance material costs.

3. EVAPORATORS

3.1. Process Description

Detailed discussion of evaporation equipment is beyond the scope of this chapter. However, thermal evaporation precedes sludge drying in several processes. Therefore, a brief description of some basic evaporators will be given here. Standard chemical engineering treatises examine evaporation in further detail (8–10).

Steam-heated evaporators are available in two major types: single-effect and multiple-effect evaporators. Single-effect evaporators are subdivided into short-tube, long-tube, and agitated-film evaporators. Here will be discussed the common, vertical, short-tube evaporator, which also constitutes the basic repeated unit in multiple-effect evaporators.

A vertical, short-tube evaporator is shown in Fig. 2. A bundle of short tubes (A), 4–8 ft long and 2–4 in. in diameter is placed in a vertical shell (B) in which the evaporating liquor is introduced. Steam condenses outside the tubes causing boiling of the liquor. The liquor spouts upward inside the tubes and returns through the downtake. Concentrated liquor is removed from the bottom of the evaporator (C) and liquid vapor is removed at (D). The cross-sectional area of the downtake is 25% of the total cross-sectional area of the tubes.

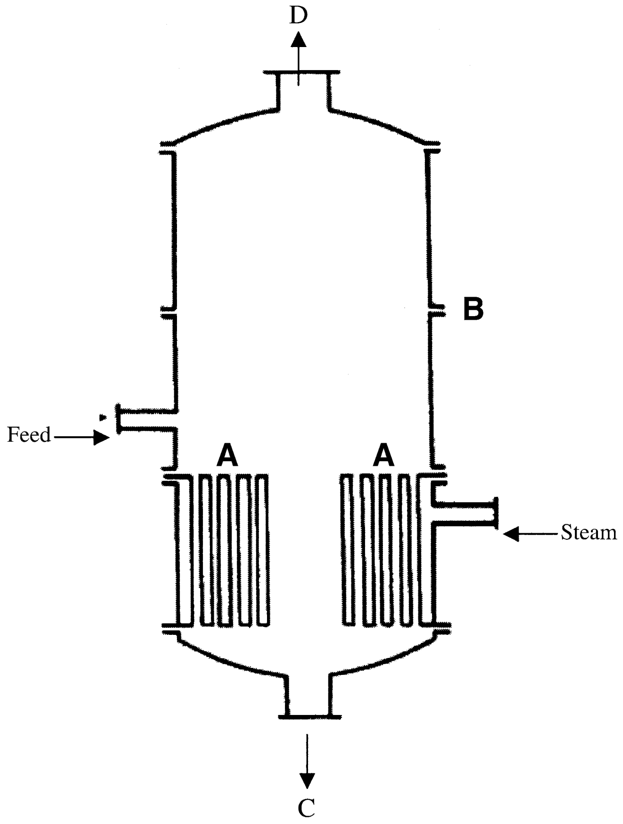

Fig. 2. A vertical short-tube evaporator. (A) Bundle of tubes; (B) shell; (C) exit of concentrated liquor; (D) vapor exit.

Short-tube evaporators are being replaced today by long-tube evaporators to achieve a higher heat transfer coefficient (1).

Single-effect evaporators can be combined in series for multiple-effect operation. Connection is arranged so that the vapor from one evaporator serves as the heating medium for the next one, as shown in Fig. 3. A vacuum is established in the last stage to remove noncondensed vapor from the system. Steam is supplied to the first stage. This arrangement results in the spreading of the pressure difference between inlet steam and final outlet condensate over all stages. The first stage operates at the highest pressure and the last one at the lowest.

Each stage operates as a single-effect evaporator with its own temperature-driving force and heat-transfer coefficient, corresponding to the pressure drop in that stage. At steady-state operation, the temperature, the concentration, and the flow rate of the feed are fixed. The inlet steam pressure and the output condensate pressure are also fixed. Operating conditions within each stage are uniquely established. The composition of the final concentrated liquor can be changed by simply adjusting the flow rate of the feed. By reducing the feed flow rate, the thick liquor concentration is increased and a new steady-state operation is reached eventually.

Evaporators often operate under vacuum to decrease the boiling point of water or solvent. This results in a larger temperature gradient between evaporating liquid and heating

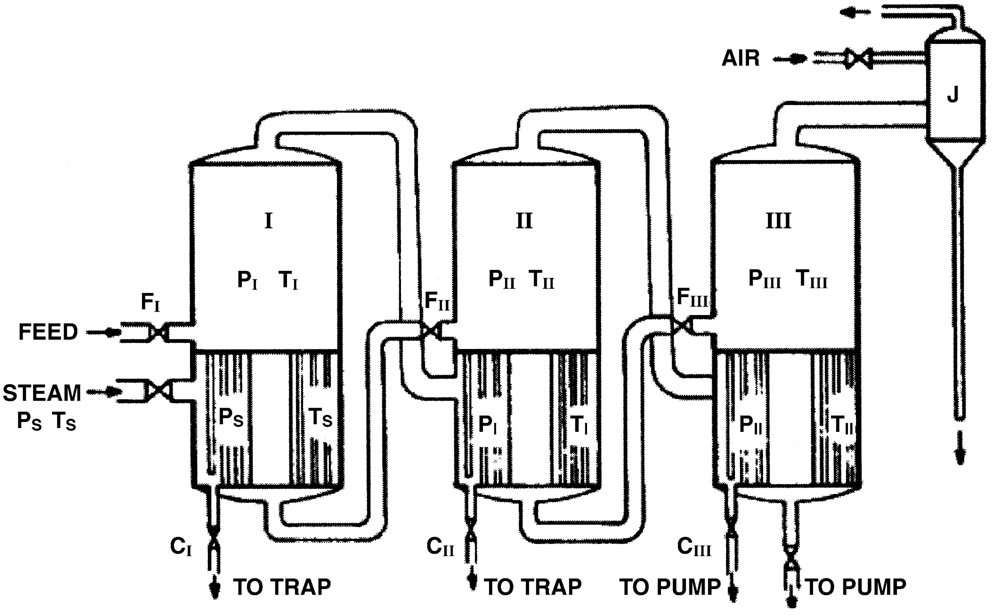

Fig. 3. Triple-effect evaporator. C_I, C_{II}, C_{III}, condensate values; F_I, F_{II}, F_{III}, feed valves; *J*, Air Injector and Condenser; *P* and *T*, pressure and temperature for each effect.

medium and in a smaller heat exchange area than if atmospheric pressure were maintained. Vacuum evaporation is important for heat-sensitive materials not only to achieve better heat transfer, but also to avoid decomposition or alteration of these materials at elevated boiling temperatures. Food and pharmaceutical slurries are, therefore, evaporated under vacuum.

Vacuum evaporation has been used in vitamin B_{12} production from wastewater sludge in Milwaukee (11). The Carver–Greenfield dehydration system uses a triple-effect evaporation step in recovering grease from municipal wastewater and industrial wastes. The Bell–Fons process uses evaporation to precipitate ferrous sulfate monohydrate and to recover sulfuric acid from the pickling liquor of steel mills (12). Similarly, calcium chloride is recovered from the industrial wastes of Columbia-Southern Chemical Company and is marketed for ice making and highway dust control. Celanese Chemical Corporation at Bishop, Texas has developed a solar evaporation process for treatment of its wastes from the production of organic chemicals.

Some typical operating data for the Carver–Greenfield process are given here. Over 65 plants exist worldwide, including an 180,000 gal/d (692 m^3/d) plant for treating 4%-solids activated sludge from the Coors Brewery and a 264,000 gal/d (1000 m^3/d) plant for a 2%-solids wastewater effluent at Hiroshima (13). Mixing sludges with oils (e.g., No. 2 fuel) helps maintain fluidity of the sludge through all stages and minimizes corrosion and scale formation in the equipment. Steam requirements have been estimated at 0.45 lb/lb H_2O at about 50 psig for a four-effect unit. Energy requirements, including steam production, are about 675 BTU/lb water, compared to 1200–2000 BTU/lb water for other dryers (2).

3.2. Process Applications and Limitations

Although the man-made evaporators are technically feasible for water evaporation, they are generally only economically feasible when the dried sludge can be sold as fertilizer or used as vitamin- and protein-enriched animal feedstock. Such possibilities will depend on the market values, as well as the attitude of the public to demand waste recycle and accept the associated environmental costs. More detailed evaporation process applications are discussed in Example 9 (Section 4.9).

Evaporation can greatly reduce the volume of wastewater requiring disposal. The water recovered from evaporation (distillate) is of high purity; therefore, the process can be used to convert waste effluent to pure or process water where other water supplies are inadequate or nonexistent. In the electroplating subcategory of the metal finishing industry, evaporation has the advantage of permitting recovery of a wide variety of plating and other process chemicals.

The evaporation process consumes relatively large amounts of energy. However, the recovery of waste heat from many industrial processes to provide a source of heat can alleviate the costs. Moreover, the equipment is sometimes highly specialized, and thus can be expensive. Another limitation is that, in some cases, pretreatment may be required to remove solids and/or bacteria that tend to cause fouling in the condenser or evaporator.

The build-up of scale on the evaporator plates reduces the heat transfer efficiency and may present a maintenance problem or increase operating cost. However, it has been demonstrated that fouling on the heat transfer surfaces can be avoided or minimized for certain dissolved solids by maintaining a seed slurry, which provides preferential sites for precipitate deposition. In addition, low-temperature differences in the evaporator will eliminate nucleate boiling and supersaturation effects.

Steam distillable impurities in the process stream are carried over with the product water and must be handled by pre- or posttreatment, if they cannot be tolerated.

Evaporator liquids, usually considered the product, must be further treated for recovery or disposal if they are not already in recoverable form. When ponding is used, the solid residues generated must also be disposed of.

3.3. Design Considerations

The evaporation process is designed on the basis of the quantity of water to be evaporated, the quantity of heat required to evaporate water from solution, and the heat-transfer rate. The necessary heat-transfer rate can be calculated on the basis of the required evaporation rate. The evaporator and operating conditions for the evaporator can then be selected to achieve the computed overall heat-transfer rate.

3.3.1. Heat Transfer

The design of evaporators depends on their required capacity and the required steam consumption. Heat is transferred from the steam in the evaporating liquid through a heating surface. The rate of heat transferred Q (BTU/h) is

$$Q = UA\Delta T \tag{1}$$

where U = the overall heat transfer coefficient, BTU/ft^2-h-°F; ΔT = the overall temperature drop between steam and evaporating liquid, °F; and A = heating surface area, ft^2.

The transferred heat raises the temperature of the liquid to its boiling point, corresponding to the absolute pressure in the evaporator, and supplies the latent heat

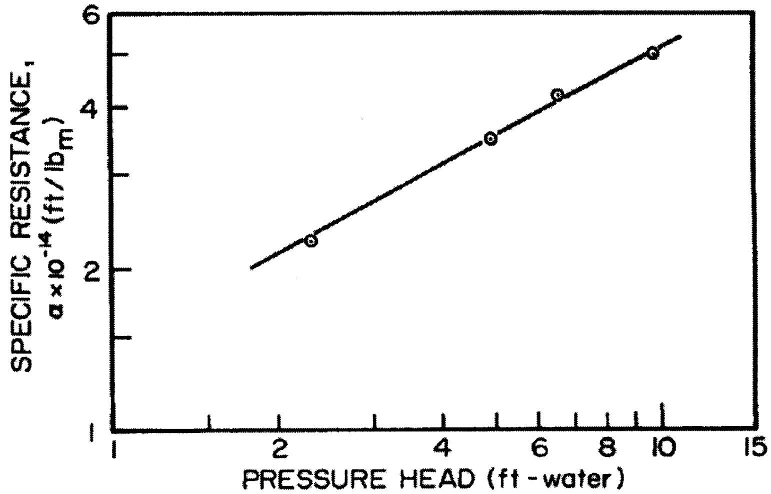

Fig. 4. Dependence of the specific resistance on pressure head.

of vaporization of water. If the feed is at a temperature above the boiling point, flash evaporation occurs.

Dissolved substances in water tend to lower the vapor pressure of water at a given temperature. Conversely, the boiling point of solutions at a given pressure is higher than that of pure water. Boiling point elevation is particularly significant for strong solutions for which Dühring's rule applies (1).

In an evaporator loaded with an appreciable depth of liquid, the boiling point increases with the depth because of the existing liquid head. Therefore, the actual boiling point is higher than that corresponding to the pressure in the evaporator, resulting in decreased capacity (8,9).

The heat-transfer coefficient expresses the facility of heat flow for a particular design and operation. The overall resistance ($1/U$) is the sum of the resistances to heat transfer on the steam side, on the liquid side, and across the tube wall:

$$Q = UA\Delta T \tag{2}$$

where D_s, D_l, and D_{ln} = the outside (steam), inside (liquid), and mean logarithmic tube diameters, respectively, ft; δ = the tube wall thickness, ft; k_T = the wall thermal conductivity, BTU-ft/ft^2·h·°F; h_l = the heat transfer coefficient on the liquid side, and h_{os} = the heat transfer coefficient on the steam side, BTU/ft^2·h·°F.

If scale forms inside and/or outside the tube walls, additional resistance terms (Fig. 4) should be added to Eq. (2). Some typical values of overall heat-transfer coefficients are given in Table 3 for various evaporator designs.

3.3.2. Heat and Material Balance

A schematic diagram of a single-effect evaporator with all streams and their properties is shown in Fig. 5. A material balance in the evaporator gives

$$F' = L' + V' \tag{3}$$

Table 3
Overall Heat Transfer Coefficients for Evaporators

Evaporator type	U (BTU/ft^2·h·°F)
1. Short tube	
a. Horizontal tube	200–400
b. Calandria type	150–500
2. Long tube, vertical	
a. Natural circulation	200–600
b. Forced circulation	400–2000
3. Coil evaporators	200–400
4. Agitated film, μ = 1cp	400

Adapted from Mc Cabe and Smith (9).

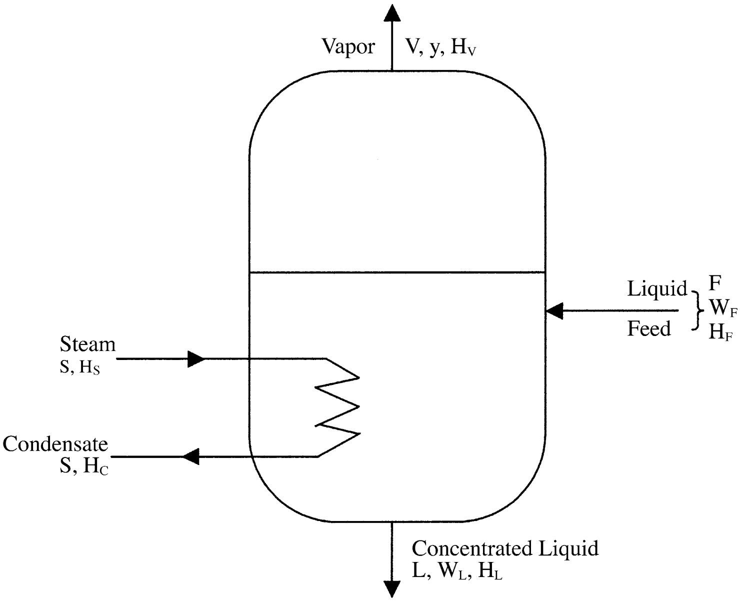

Fig. 5. Schematic diagram of a single-effect evaporator.

where F' = the weight of feed, lb/h; L' = the weight of the resulting thick liquor, lb/h; and V' the weight of the vapor phase, lb/h.

If w_F = the weight fraction of water in the feed, w_L = the weight fraction of water in the concentrated liquid, and y = the weight fraction of evaporated water

A material balance for water yields

$$w_F F' = w_L L' + yV' \tag{4}$$

The heat supplied by the steam is

$$Q_S = S'(H_S - H_C) \tag{5}$$

where Q_s = rate of heat transfer by steam, BTU/h; S' = the weight of steam supplied to the evaporator lb/h; and H_s and H_c = the enthalpies of steam and condensate, respectively, BTU/lb. Complete condensation of saturated steam is assumed with no condensate cooling in the evaporator.

A heat balance for the condensing liquid gives

$$Q_L = (\text{Heat out, in vapor and thick liquid}) - (\text{Heat in, in feed}) \tag{6}$$

$$Q_L = (L'H_L + V'H_V) - F'H_F \tag{7}$$

where Q_L = rate of heat transfer to liquid, BTU/h; H_F = enthalpy of feed, BTU/lb; H_L = enthalpy of thick liquid, BTU/lb; and H_V = enthalpy of vapor, BTU/lb.

At steady state, $Q_L = Q_S$ and using Eq. (3),

$$S'(H_S - H_C) = L'(H_L - H_F) + V'(H_V - H_F) \tag{8}$$

Normally, low-pressure steam is used in evaporation. Although high-pressure steam could provide a larger temperature gradient across the heating surface for given conditions, such steam is usually valuable for energy generation.

3.3.3. Multiple-Effect Evaporators

Equation (1) holds for heat transfer in each stage of a multiple-effect evaporator (Fig. 3). For the first stage (I) then

$$Q_{\mathrm{I}} = U_{\mathrm{I}} A_{\mathrm{I}} \Delta T_{\mathrm{I}} \tag{9}$$

If the feed is at or close to the boiling point corresponding to the conditions in the first stage, essentially all Q_{I} goes into vaporizing water in this stage. At steady state, this water vapor will condense around the tubes of the second stage to vaporize an almost equal amount of water in stage II. The condensate in this stage is at about the same temperature as the vapor of the boiling liquid in the first stage.

The heat exchanged in stage II is

$$Q_{\mathrm{II}} = U_{\mathrm{II}} A_{\mathrm{II}} \Delta T_{\mathrm{II}} \tag{10}$$

From the operation of this stage it follows that the amounts of heat exchanged in effects II and I are almost equal.

$$U_{\mathrm{I}} A_{\mathrm{I}} \Delta T_{\mathrm{I}} = U_{\mathrm{II}} A_{\mathrm{II}} \Delta T_{\mathrm{II}} \tag{11}$$

By similar reasoning, the same amount of heat is exchanged in the third effect, thus

$$U_i A_i \Delta T_i = Q_i = Q \tag{12}$$

where Q = a constant and the subscript denotes the effect.

Usually, the heating surface areas of all effects are equal for construction economy, therefore,

$$\frac{1}{U} = \frac{D_s}{D_l h_l} + \frac{\delta}{k_T} \cdot \frac{D_s}{D_{ln}} + \frac{1}{h_{os}} \tag{13}$$

Equation (13) suggests that the temperature drops are inversely proportional to the overall heat-transfer coefficients in each effect, e.g.,

$$F' = L' + V' \tag{14}$$

It should be emphasized that Eqs. (12)–(14) are only approximate. The total heat exchanged in an *N*-effect evaporator is the sum of all Q_i:

$$w_F F' = w_L L' + yV' \tag{15}$$

If all A_i and U_i are equal to a constant value *A* and *U*, respectively, then

$$Q_S = S'(H_S - H_C) \tag{16}$$

where ΔT_T = the total temperature drop across the system.

Equation (15) suggests that the total heat exchanged in the multiple effect evaporators would be the same as the heat exchanged in a single effect evaporator with the same *U* and *A* as each effect, operating under a temperature gradient ΔT_T. Therefore, the capacity of a multiple-effect evaporator is no better than that of an equivalent single-effect unit. Here, capacity is defined as the total rate of water vaporization (lb/h). However, significant steam economy is achieved. Each pound of steam supplied to an *N*-effect evaporator vaporizes approximately N pounds of water. In a single-effect evaporator, each pound of steam vaporizes only about 1 lb of water. To arrive at these approximate relationships, liquid heating and any heat losses have been neglected.

4. DESIGN EXAMPLES

4.1. Example 1

Sludge having 8% solids is concentrated in a single-effect evaporator to 35% solids. The evaporator operates at 0.95 psia and uses steam at 10 psig. If the feed rate is 10,000 lb/h at 70°F, and the heat-transfer coefficient is 400 BTU/ft^2·h·°F, calculate: (a) the amount of steam required; (b) the amount of water evaporated per pound of steam (economy); and (c) the heating surface area. Neglect the boiling point elevation and heat of dilution for the sludge. The specific heat capacity of the feed sludge is $C_{P,F}$ = 0.88 BTU/lb·°F.

Solution:

Material Balance:

Water in feed: $Q_L = (\textit{Heat out } [\textit{in vapor and thick liquid}]) - (\textit{Heat in } [\textit{in feed}])$

Water in concentrate: $Q_L = (L'H_L + V'H_V) - F'H_F$

Water evaporated: = 9.65 lb water/lb solids

For a feed F' = 10,000 lb sludge/h, then,

V' = 10,000 × 0.08 × 9.65 = 7720 lb water/h

L' = 10,000 – 7720 = 2380 lb liquor/h

Steam Requirements: The steam requirements are estimated from Eqs. (3)–(7) rearranged to give

$$S'\Delta H_v = (F' - L')H_v + L'H_L - F'H_F \tag{17}$$

where ΔH_v is the latent heat of vaporization of steam at 10 psig.
If the temperature of the concentrated liquor, T_L(°F) , is considered as a reference temperature (T_R, °F) Since $T_L = T_R$, then

$$H_L = C_{P,L}\,(T_L - T_R) = 0 \tag{18}$$

and

$$H_F = C_{P,F}\,(T_F - T_R) = C_{P,F}\,(T_F - T_L) \tag{19}$$

where T_F = temperature of feed sludge, °F; $C_{P,L}$ = specific heat capacity of the concentrated liquor, BTU/lb·°F.

The enthalpy of the vapor H_v with respect to the thick liquor now represents the latent heat of vaporization at the operating pressure in the evaporator. Equation (18) then yields

$$S'\Delta H_v = (F' - L')H_v - F'C_{P,F}\,(T_F - T_L) \tag{20}$$

From steam tables (such as Appendix), at 10 psig, the temperature of steam (T_S) can be found:

$$T_s = 239.4°\text{F}$$

$$\Delta H_v = H_s - H_c = 952.6 \text{ BTU/lb}$$

At 0.95 psia,

$$T_L = 100°\text{F}$$

$$H_v = 1037 \text{ BTU/lb}$$

With the feed entering at T_F = 70°F, the steam rate is

$$S'(H_S - H_C) = L'(H_L - H_F) + V'(H_V - H_F)$$

Steam Economy

$$\frac{V'}{S'} = \frac{7720}{8572} = 0.90 \text{ lb water evaporated/lb steam}$$

Heating Surface Area

$$A = \frac{Q}{U(T_S - T_L)} = \frac{S'\Delta H_v}{U(T_S - T_L)} \tag{21}$$

or

$$A = \frac{8{,}572 \times 952.6}{400(239.4 - 100)} = 146.4 \text{ft}^2$$

4.2. Example 2

Sludge is concentrated in a triple-effect evaporator to recover vitamin B_{12}. The first effect operates at 8 psig, whereas the last effect operates at a temperature of 110°F. If the overall heat transfer coefficients are 400, 320, and 240 for the first, second, and third effects, respectively, estimate the liquor boiling temperature in each stage. Neglect any boiling point elevation.

Solution:

Assuming that all effects have the same heating surface area, Eqs. (13) and (14) give

$$\Delta T_{\mathrm{I}} : \Delta T_{\mathrm{II}} : \Delta T_{\mathrm{III}} : (\Delta T_{\mathrm{I}} + \Delta T_{\mathrm{II}} + \Delta T_{\mathrm{III}}) = 1/U_{\mathrm{I}} : 1/U_{\mathrm{II}} : 1/U_{\mathrm{III}} : (1/U_{\mathrm{I}} + 1/U_{\mathrm{II}} + 1/U_{\mathrm{III}}) \quad (22)$$

Thus

$$U_i A_i \, \Delta T_i = Q_i = Q \tag{23}$$

where

$$U_i \Delta T_i = \frac{Q}{A} = \text{constant}$$

From steam tables, at 8 psig

$$T_{\mathrm{I}} = 235°\mathrm{F}$$

and

$$\Delta T_T = 235 - 110 = 125°\mathrm{F}$$

Thus

$$\Delta T_{\mathrm{I}} = \frac{125}{\left(\frac{1}{400} + \frac{1}{320} + \frac{1}{240}\right)} \left(\frac{1}{400}\right) = 32°\mathrm{F}$$

Similarly,

$$\Delta T_{\mathrm{II}} = \frac{125}{\left(\frac{1}{400} + \frac{1}{320} + \frac{1}{240}\right)} \left(\frac{1}{320}\right) = 39.8°\mathrm{F}$$

and

$$\Delta T_{\mathrm{III}} = \frac{125}{\left(\frac{1}{400} + \frac{1}{320} + \frac{1}{240}\right)} \left(\frac{1}{240}\right) = 53.2°\mathrm{F}$$

from which

$$\begin{aligned} T_{\mathrm{II}} &= T_{\mathrm{I}} - \Delta T_{\mathrm{I}} \\ &= 235 - 32 = 203°\mathrm{F} \end{aligned}$$

and

$$T_{\mathrm{III}} = T_{\mathrm{II}} - \Delta T_{\mathrm{II}} = 203 - 39.8 = 163.2°\mathrm{F}$$

4.3. Example 3

Analyze the performance of a 60.5-ft diameter (max. surface area on top) evaporation lagoon in Spokane County, Washington, USA, for treating 5400–5580 gal per month of liquid sludge in April–June, assuming the following data are known:

Month	April	May	June
Average flow, gal	5400	5580	5400
Average evaporation, in.	5.54	7.79	9.26
Average precipitation, in.	1.00	1.00	1.20

Solution:

Based on the above data, the water mass balance analysis has been conducted, and its results are:

Month	April	May	June
Net Evaporation, in.	4.54	6.79	8.06
Net Evaporation, gal/ft^2	2.83	4.23	5.02
Evaporation Area, ft^2	1908	1319	1076
Net Evaporation, gal	5400	5580	5400

(Note: 1 in. = 0.6233 gal/ft^2)

4.4. Example 4

Discuss the following for the sludge evaporation lagoon in Example 3:

(a) Structural elements.
(b) Performance expectations.
(c) A cable and scraper system for the sludge evaporation lagoon.
(d) Operation and maintenance.

Solution:

(a) Structural Elements: The retaining walls for drying lagoons are typically earthen dikes 0.7–1.4 m (2–4 ft) high with a side slope of 1:3. Although the lagoon is typically rectangular in shape to facilitate sludge removal, a circular 60.5-ft diameter evaporation lagoon is technically feasible for efficient water evaporation. Figure 1 shows that the required equipment includes sludge feed lines and pumps, supernatant decant lines, and sludge removal equipment. The last may include trucks, front-end loaders, bulldozers, or draglines, depending on the size of the operation.

(b) Performance Expectations: Solids concentrations in the range of 15–40% are expected in the sludge removed from the lagoon, although concentrations can be higher in arid climates. These lagoons share a common problem with other air drying processes in that a surface crust forms early in the evaporative stage, which then restricts further evaporative water losses. This problem is minimized with the paved drying beds that use mechanical equipment to move around the bed to turn and mix the sludge. Similar equipment and procedures can be used in drying lagoons, if the depth of sludge permits. Floating devices can also be used.

(c) Larger scale facilities may use cable and scraper system as shown in Fig. 6.

(d) Operation and Maintenance: The routine operational activities consist of sequential sludge applications and decantations until the lagoon contains the design volume of sludge. The periodic break-up or removal of the surface crust then ensures continued evaporation. Sludge removal is labor intensive but occurs infrequently. Maintenance activities include care of equipment and dikes and control of dike vegetation. Some sludge drying lagoons may require insect and odor control. The labor requirements for sludge drying lagoons are shown in Table 2.

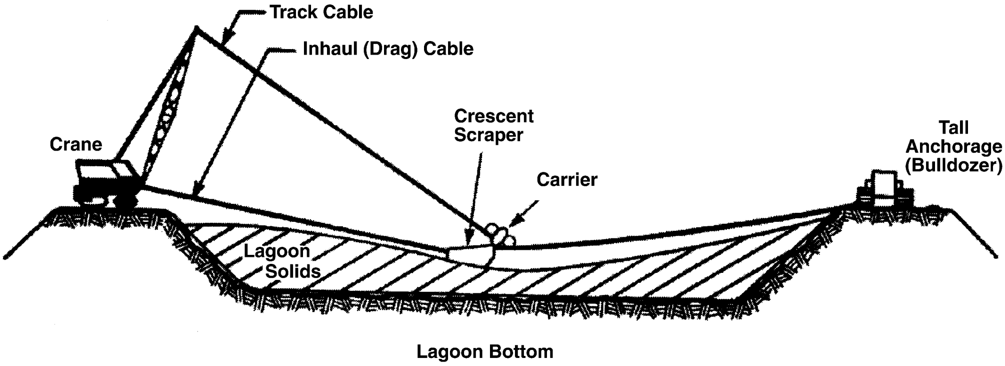

Fig. 6. Cable and scraper system for sludge drying lagoons.

4.5. *Example 5*

Obtain the annual evaporation data for the United States. Discuss its applicability for a sludge evaporation lagoon design.

Solution:

Annual evaporation data for the contiguous United States are presented in Fig. 7, which was compiled by Buonicore and Davis (7). The compiled annual evaporation data will be suitable for long-term operation of a sludge evaporation lagoon. For better lagoon design and operation, the environmental engineer in charge of design and operation should obtain the local 10-yr monthly evaporation and precipitation data and perform calculations similar to Example 3.

4.6. *Example 6*

Introduce the manufacturers of commercial prefabricated evaporators.

Solution:

The manufacturers of commercial prefabricated evaporators can be found in refs. 14–18.

4.7. *Example 7*

Briefly define "evaporation process," discuss the process energy source, and explain the difference between "evaporation" and "drying" from technical viewpoints.

Solution:

Although the sludge evaporation lagoon process is also called a sludge drying process, there is a difference between evaporation and drying.

Evaporation is a concentration process involving removal of water from a solution by vaporization to produce a concentrated residual solution. The energy source may be synthetic (steam, hot gases, and electricity) or natural (solar or geothermal). The process offers the possibility of total wastewater elimination with only the remaining concentrated solution requiring disposal and also offers the possibility of recovery and recycle of useful chemicals from wastewater.

Evaporation differs from drying in that the evaporation residue is usually a highly viscous liquid, and the vapor a single component. When the vapor is a mixture, no attempt normally is made in the evaporation step to separate the vapor into different components.

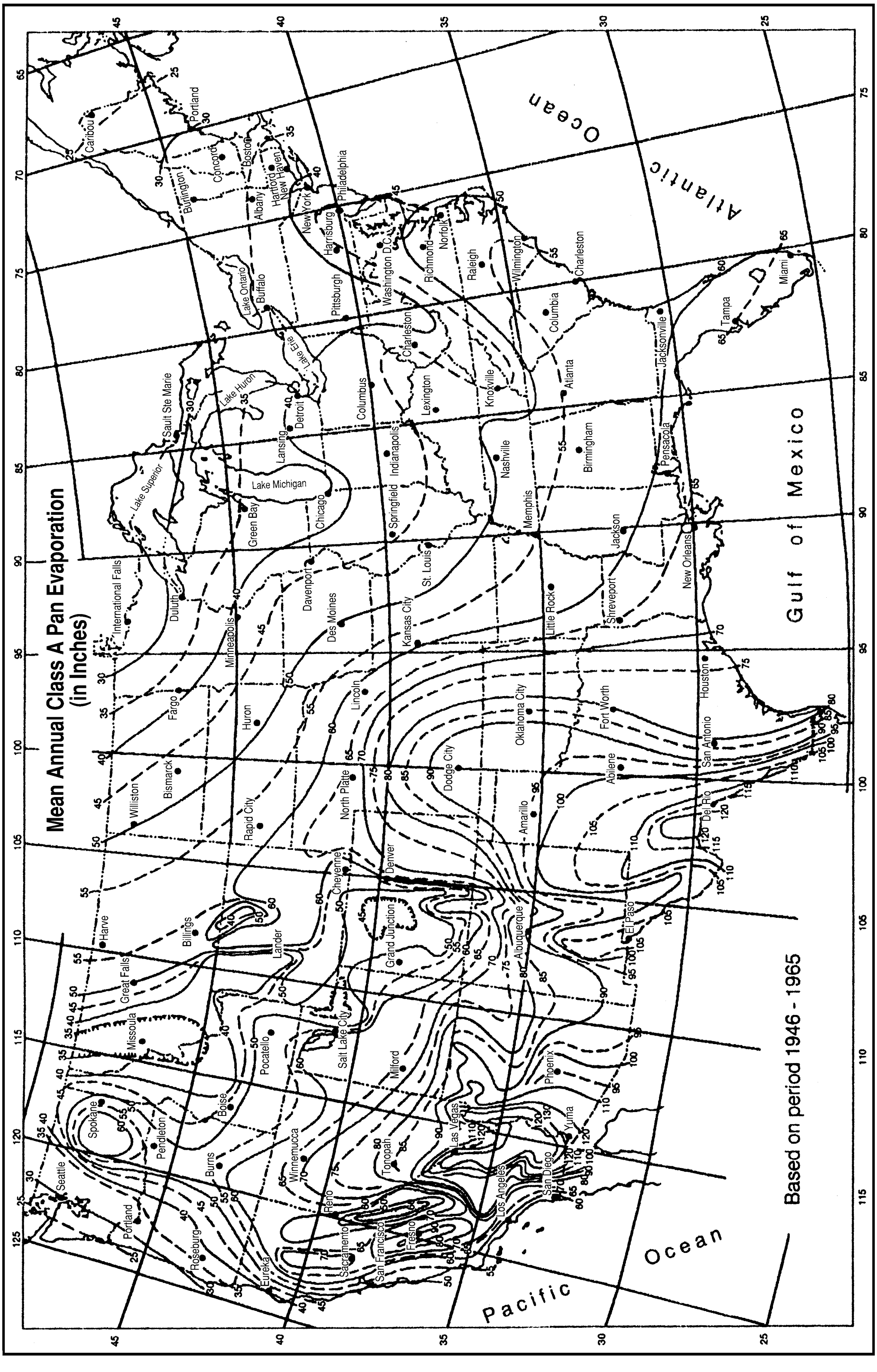

Fig. 7. Annual evaporation data for the contiguous United States.

4.8. Example 8

There are many types and modifications existing for the evaporation process. The text divides the evaporation process into natural evaporation and man-made evaporation. Please divide the evaporation process into other categories based on energy source and mechanical operation.

Solution:

The evaporation process can be divided into the broad categories of steam evaporation and solar evaporation.

1. Steam Evaporation: In this process, steam is used to raise the temperature of solution to its boiling point. The process is carried out either at a pressure less than atmospheric (vacuum evaporation) or at atmospheric pressure (atmospheric evaporation).
   (a) Vacuum Evaporation: In this modification, the pressure is lowered to cause the liquid to boil at a reduced temperature and to protect any organic fraction of the evaporating solution from thermal decomposition. All of the water vapor is condensed and, to maintain the vacuum condition, noncondensible gases (air in particular) are removed by a vacuum pump. Vacuum evaporation may be either single or multiple effects. For example, in double-effect evaporation, the water vapor from the first evaporator is used to supply heat to a second evaporator operated at a lower pressure. Roughly equal quantities of wastewater are removed in each evaporator; thus, the double effect system removes twice the water of a single-effect system, at nearly the same cost in energy but with added capital cost and complexity. Thermal or mechanical vapor recompression is another energy-conservation technique available, which enables heat transfer from the condensing water vapor to the evaporating wastewater.
   Vacuum evaporating equipment may be classified as submerged tube or rising (climbing) film. A brief description of the two follows:
      (i) Submerged tube (In most commonly used submerged tube evaporators, the heating and condensing coils are contained in a single vessel to reduce capital cost. The vacuum in the vessel is maintained by an eductor-type pump, which creates the required vacuum by the flow of the condenser cooling water through a Venturi. Wastewater accumulates in the bottom of the vessel and is evaporated by means of submerged steam coils. The resulting water vapor condenses as it contacts the condensing coils in the top of the vessel. The condensate then drips off the condensing coils into a collection trough that carries it out of the vessel. Concentrate is removed from the bottom of the vessel.
      (ii) Rising film (The major elements of the rising film evaporator are the evaporator, the separator, the condenser, and the vacuum pump. Wastewater is “drawn” into the system by the vacuum so that a constant liquid level is maintained in the separator. Liquid from the separator enters the steam-jacketed evaporator tubes and is partially evaporated. A mixture of vapor and liquid returns to the separator, with the liquid removed by mesh entrainment and continuously circulated from the separator back to the evaporator. The vapor entering the separator flows into the condenser where it is condensed as it flows down through the condenser tubes. The condensate, along with any entrained air, is pumped out of the bottom of the condenser by a liquid ring vacuum pump. Thus, the liquid seal provided by the condensate keeps the vacuum in the system from being broken.

   (b) Atmospheric Evaporation: Atmospheric evaporators do not recover the distillate for reuse and do not operate under a vacuum. Wastewater is evaporated by using it to humidify air flowing through a packed tower. The humidified air is exhausted to the atmosphere, eliminating the need for a condenser.

2. Solar Evaporation: Natural evaporation from wastewater impoundments located in arid regions is a technique practiced at many operations to reduce discharges to zero or nearly zero. Successful implementation depends on favorable climatic conditions (not evaporation) and on the availability of land. Land requirements can be significant in areas where the net evaporation value is small, and a large surface area of water must be exposed. In some instances where impoundment is not practical for the total wastewater discharge, impoundment of smaller, highly contaminated wastewaters from specific processes may afford significant advantages.

Solar evaporation can be substantially increased by a variety of techniques that mechanically improve mass transfer rates, such as spraying. The wastewater is sprayed under pressure through nozzles producing fine aerosols, which are evaporated in the atmosphere. The driving force for this evaporation is the difference in relative humidity between the atmosphere and the humidity within the spray area. Temperature, wind speed, spray nozzle height, and pressure are all variables that affect the amount of wastewater that can be evaporated.

4.9. Example 9

Evaporation is a well-defined and well-established process. The technology is proven and its application is expanding. Evaporation is very reliable and generally does not require extensive operator attention. This chapter's text places emphasis on sludge evaporation/drying. Please explain other environmental applications of the evaporation process.

Solution:

Evaporation can be used for a variety of purposes including dehydration, recovery, separation, and concentration. Evaporation is especially useful in the treatment and disposal of specific high-strength, low volume process waste streams. The following industries use evaporation methods on a widespread basis for waste treatment and/or recovery of chemicals:

(a) Metal Finishing
(b) Explosives Manufacturing
(c) Timber Products Processing

The following industries use evaporation on a limited basis:

(a) Inorganic Chemicals Manufacturing
(b) Aluminum Forming
(c) Battery Manufacturing
(d) Pharmaceutical Manufacturing
(e) Nonferrous Metals Manufacturing
(f) Organic Chemicals Manufacturing
(g) Paint and Ink Formulation
(h) Petroleum Refining
(i) Rubber Processing
(j) Textile Mills

In the metal finishing industry, evaporation is a common technology for recovery of plating chemicals from rinse water. Evaporation achieves recovery by distilling the wastewater until there is sufficient concentration of plating chemicals to allow reuse in the plating operation. The water vapor is condensed and returned to the rinse tank. In the pulp and paper industry, evaporation is used to concentrate the spent liquor into a viscous mass called "strong black liquor." The strong black liquor is then burned to recover heat and chemicals.

Table 4
Investigation of Sludge Production, Lenox Water Treatment Plant, November, 1982

Date	Influent flow (gpm)	Sludge flow (gpm)	Sludge TSS (mg/L)
11/01/82	429–545	3	3271
11/02/82	390–545	2	1334
11/03/82	390–519	2	2611
11/04/82	519	4	3051
11/05/82	490–519	5	1771
11/06/82	391–519	5	1771
11/07/82	391–519	6	1771
11/08/82	391–535	3	1771
11/09/82	391–535	4	1029
11/10/82	391	5	1029
11/11/82	391–535	2	1440
11/12/82	535–536	2	2555
11/13/82	391–536	4	3180
11/14/82	536	2	1035
11/15/82	391–536	5	3810
11/16/82	391–527	4	948
11/17/82	527	4	488
11/18/82	387–527	4	3000
11/19/82	387–536	3	1303
11/20/82	536	3	1548
11/21/82	387–547	3	1548
11/22/82	547	3	1258
11/23/82	388–547	8	2732
11/24/82	388–595	3	983
11/25/82	477–481	5	1346
11/26/82	476–481	7	4621
11/27/82	471–476	3	7705
11/28/82	471–500	1	12,535
11/29/82	479–500	2	3211
11/30/82	479	2	3211
Range	387–595	1–8	488–12,535
Average		3.63	2596

Adapted from Krofta and Wang (30,31).

4.10. *Example 10*

An innovative potable filtration plant with a design capacity of 1.2 MGD has been reliably serving 10,000 residents and tourists in the town of Lenox, Massachusetts, USA, since July 1982. Its process system consists of chemical flocculation, dissolved air flotation, and automatic backwash and sand filtration. It substantially improves upon the conventional flocculation, sedimentation, and filtration system in performance, capability, operation, maintenance, and energy use (19–42).

The detention time of the Lenox flotation system including flocculation, flotation, filtration, and clear-well is only 15 min in comparison with the conventional system's 6–9 h

Table 5
Sludge Generation at Lenox Water Treatment Plant

Parameters	Data
Average plant flow	
gpm	521.0
MGD	0.75
Peak plant flow	
gpm	694.4
MGD	1.0
Raw sludge concentration	
mg/L	2600.0
Sludge production rate	
dry lb/d/MGD	150.8
Average sludge production	
dry lb/d/MGD	113.0
Peak sludge production	
dry lb/d/MGD	150.8
Sludge flow rate	
gpm/MGD	4.84
% influent flow	0.70
Average sludge flow	
gpm	3.63
gph	217.8
Peak sludge flow	
gpm	4.84
gph	290.4

Adapted from Krofta and Wang (30,31). 1 gpm = 1 gallon per minute = 3.785 liters per minute; 1 gph = 1 gallon per hour = 3.785 liters per hour; 1 MGD = 1 million gallons per day = 3.785 million liters per day; 1 lb = 454 grams.

Table 6
Chemical Treatment Summary

Month in 1982	Water treated (gal)	Alum dosage (mg/L Al_2O_3)	Alum dosage (lb)	Alum residue (mg/L Al_2O_3)	Polymer dosage (mg/L)	Polymer dosage (lb)	Polymer residue (mg/L)
July	11,622,900	1.99	192.60	0.31	0	0	0
August	31,480,646	2.75	723.27	0.50	0.46	120.25	0.09
September	20,461,473	2.90	491.75	0.43	0.02	4.12	0
October	24,471,287	2.27	463.52	0.37	0.11	22.46	0.004
November	20,351,372	2.40	407.76	0.35	0.84	142.16	0.03
December	21,113,800	3.39	597.33	0.32	0.44	77.17	0.07
Total	129,501,478		2876.23			366.16	
(Average)		(2.66)		(0.38)	(0.34)		(0.03)

Adapted from Krofta and Wang (30,31) and Wang (41,42).
1 gal = 3.785 liters; 1 lb = 454 grams.

Table 7
Analytical Data of Composite Settled Lagoon Sludge*, Lenox Water Treatment Plant, Lenox, Massachusetts

Parameter	Sludge data	US EPA limits for land application
pH, unit	6.9	None
Total suspended solids, mg/L**	30,425	None
Volatile suspended solids, mg/L**	3420	None
Fixed suspended solids, mg/L	27,005	None
Cadmium, mg/kg dry sludge	<0.14	16
Chromium, mg/kg dry sludge	54	140
Lead, mg/kg dry sludge	25	500
Copper, mg/kg dry sludge	64	850
Nickel, mg/kg dry sludge	80	82
Zinc, mg/kg dry sludge	14	1740
Aluminum, mg/kg dry sludge	30,500	None
Iron, mg/kg dry sludge	NA	None
Mercury, mg/kg dry sludge	BD	5

Adapted from Krofta and Wang (30,31) and Wang (41,42).
*The settled lagoon sludge was accumulated in the period from May 21 to Nov. 21, 1982.
**Total suspended solids and volatile suspended solids are average values of eight sludge samples.
NA = Not available.
BD = Below detection limit of AA.

of detention time. Because of the Lenox flotation plant's compact design (diameter = 22 ft; depth = 6 ft), its equipment installation cost, housing cost, heating cost, land requirement, etc., are all significantly reduced.

The Lenox Water Treatment Plant has been using a sludge evaporation lagoon (sludge drying lagoon) for thickening its alum sludge. Introduce the plant's sludge generation data and its sludge evaporation lagoon's performance.

Solution

(a) Sludge Generation: An investigation of sludge production at the Lenox flotation plant was conducted in November, 1982, and the results are presented in Tables 4 and 5. Table 6 documents the chemical consumption of the Lenox plant in the sludge-testing period. It can be seen that the plant's chemical consumption was much lower than that of a comparable conventional water purification plant. However, the sludge production rate was estimated to be 150.8 dry lb/d/MGD, as shown in Table 5.

An engineer's rule of thumb for sludge production rate of a comparable conventional water purification plant is usually set at 75 dry lb/d/MGD. It is believed that the low sludge production rate of a conventional plant is caused by discharge of filter backwash wastewater, without recycle. Thus, the sludges in the discharged wastewater are not included in sludge quantity estimation (30,31,44):

It is understandable that the potable water flotation plant recycles its backwash wastewater for reuse, and in turn, has higher sludge production rate (150.8 instead of 75 dry lb/d/MGD) because almost all sludges are captured by dissolved air flotation. Table 5 further confirms the plant's sludge flow rate is about 0.7% of influent flow rate. The raw sludge concentration of TSS is about 2600 mg/L.

Table 8
Sludge Evaporation and Monitoring at Lenox Water Treatment Plant

Date	Lagoon flow (gpm)	Lagoon and filter effluent: Turbidity (NTU)	TSS (mg/L)	Color (unit)
11/04/82	3.8–4.0	1.4	–	4
11/07/82	5.4–6.0	2.6	2.4	5
11/08/82	3.0–5.6	1.8	0.9	5
11/09/82	4.0–5.3	2.7	–	5
11/20/82	5.0–6.2	3.0	3.9	6

Adapted from Krofta and Wang (30,31). Fresh raw alum sludge (non-dewatered) was discharged directly into the sludge lagoon; 1 gpm = 1 gallon per minute = 3.785 liters per minute.

(b) Sludge Evaporation Performance: The data documented in Table 7 are for sludge handling and disposal (30,31,41,42). For freewheeling automatic operation without an operator's attention, the average sludge flow and sludge concentration (TSS) were 3.63 gpm and 2600 mg/L, respectively. By manual operation, with the operator's attention on June 29, 1982, the sludge concentration was as high as 15,800 mg/L, and the sludge flow was as low as 0.3 gpm. The floated sludge was discharged into a sludge lagoon with a built-in slow sand filter for disposal. The analytical data in Table 5 are for the settled lagoon sludge accumulated in the period from May 21 to November 21, 1982. It can be seen that the settled lagoon sludge with a consistency of about 3% meets the US Environmental Protection Agency limits for land application. The lagoon sludge contained mainly inorganic fixed suspended solids (27,005 mg/L), or, more specifically, the non-toxic aluminum (30,500 mg/kg dry sludge). All heavy metal contents were extremely low. The lagoon overflow passed though a slow sand filter, and eventually reached a small creek. The November 1982 lagoon operational data in Table 8 show that the slow sand filter effluent was as clean as reservoir raw water. When there is a water shortage, the lagoon/filter effluent can be pumped back to the Lower Root Reservoir for reuse, so every drop of water can be conserved.

A Discharge Permit for discharging the lagoon effluent from the Lenox Water Treatment Plant to a nearby stream has been granted by the Commonwealth of Massachusetts, Department of Environmental Quality. All dewatered alum sludge from the Lenox Water Treatment Plant is discharged to the town's Wastewater Treatment Plant.

The following paragraphs introduce the physical structures of the sludge evaporation lagoon system.

The Lenox plant's sludge evaporation lagoon consisting of a sludge lagoon and a slow sand filter was designed for holding and thickening of an average sludge flow of 3.63 gpm.

The lagoon's inlet and outlet are located at opposite ends. Its size is approx 31 ft W × 47.5 ft L at the top and 18.75 ft W × 42.5 ft L at the bottom, with a side slope of 1 1/2 to 1. Its depth is about 6 ft. The lagoon effluent is discharged to the slow sand filter via a spillway on a dividing concrete wall between the lagoon and the filter. The lagoon overflow rate and weir overflow rate are less than 500 gpd/ft^2 and less than 2000 gpd/ft, respectively.

The slow sand filter has a dimension of approx 34 ft L × 20 ft W, and is packed with 2 ft of coarse sand (0.5–0.7 mm effective size), 3 in. of small-diameter gravel under the sand, and 9 in. of graded gravel under the small diameter gravel. Its loading rate is equal to or less than 15 gpd/ft^2. The slow sand filter further polishes the lagoon effluent. The filter effluent is as clean as the reservoir raw water.

The entire sludge evaporation lagoon system has been constructed so as to provide for cleaning without interference with normal operation.

The lowest elevation of the sludge evaporation lagoon has been kept above ground water level to avoid being overflowed with ground water.

A similar but more improved evaporation lagoon system has been used at Feura Bush Filtration Plant of the City of Albany, New York, USA. The City of Albany's evaporation lagoon system involves the use of natural evaporation, freezing, thawing, and sedimentation processes for treatment of combined filter backwash water and sedimentation waste sludge.

NOMENCLATURE

A	Exposed or cross-sectional area, ft^2
$C_{P,F}$	Specific heat capacity of feed sludge, BTU/lb·°F
$C_{P,L}$	Specific heat capacity of thick liquor, BTU/lb·°F
D_i	Inside tube diameter, ft
$D_{\ln}$	Mean logarithmic diameter, ft
D_s	Outside tube diameter, ft
F'	Mass of mixture, lb/h
H_c	Enthalpy of condensate, BTU/lb
H_F	Enthalpy of feed sludge, BTU/lb
H_L	Enthalpy of thick liquor, BTU/lb
H_S	Enthalpy of steam, BTU/lb
H_V	Enthalpy of vapor, BTU/lb
h_i	Heat transfer coefficient, BTU/ft^2·h·°F
h_{os}	Heat transfer coefficient on steam side, BTU/ft^2·h·°F)
ΔH_V	Latent heat of vaporization of steam, BTU/lb
k_T	Thermal conductivity of wall, BTU ft/ft^2·hr·°F
L'	Mass of concentrated liquid, lb/h
Q	Rate of heat transfer, BTU/h
Q_L	Rate of heat transfer to liquid, BTU/h
Q_S	Rate of heat transfer by steam, BTU/h
S'	Mass of steam, lb/h
T_L	Temperature of concentrated liquid, °F
T_R	Reference temperature, °F
T_S	Steam temperature, °F
ΔT	Overall temperature drop, °F
U	Overall heat transfer coefficient, BTU/ft^2·h·°F
V'	Mass of vapor phase as water, lb/h
w_F	Weight fraction of water in the feed
w_{L}	Weight fraction of water in the concentrated liquid
W_s	Weight of solids, lb
y	Weight fraction of evaporated vapor
δ	Tube wall thickness, ft

Subscripts

a	Air
I, II, III	Interface; stage of evaporators
F	Feed
L	Thick liquor
R	Reference
V	Vapor

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Appendix: Steam Tables (45)

Temperature (°F)	Pressure (psia)	ΔH_v (BTU/lb)	H_v (BTU/lb)
100	0.95[1]	1036.98	1105.02
230	20.781	958.81	1157.12
240	24.968[2]	952.27	1160.70

[1] 0.95 psia corresponds to a T_L of 100°F, with $\Delta H_v = 1037$ BTU/lb.

[2] 10 psig corresponds to 24.696 psia. Interpolation yields a T_S of 239.4°F and 952.6 BTU/lb. 8 psig corresponds to 22.696 psia. Interpolation yields a T_L of 235°F.

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High Temperature Thermal Processes

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CONTENTS

1. INTRODUCTION

Since the early 1900s, high temperature processes have been used for combustion of municipal wastewater solids. The popularity of these processes has fluctuated greatly because of their adaptation from the industrial combustion field. In the past, combustion of wastewater solids was both practical and inexpensive. Solids were easily dewatered, and the fuel required for combustion was cheap and plentiful. In addition, air emission standards were virtually nonexistent.

In today's environment, wastewater solids are more complex and include sludges from secondary and advanced waste treatment processes. These sludges are more difficult to dewater, and thereby increase fuel requirements for combustion. Environmental concerns with air quality, and increased energy cost, are significant factors in the feasibility of high temperature processes for combustion of municipal solids.

However, recent developments in more efficient solids dewatering processes and advances in combustion technology have renewed an interest in the use of high temperature processes for specific applications. High temperature processes should be considered where available land is scarce, stringent requirements for land disposal exist, destruction of toxic materials is required, or the potential for recovery of energy exists either with wastewater solids alone or combined with municipal refuse.

High temperature processes have several advantages and disadvantages compared with other methods (1). The three main advantages are as follows:

a. **Maximum volume reduction.** Reduces volume and weight of wet sludge cake by approx 95%, thereby reducing disposal requirements.

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b. **Detoxification.** Destroys or reduces toxics that may otherwise create adverse environmental impacts (2).
c. **Energy recovery.** Potentially recovers energy through the combustion of waste products, thereby reducing the overall expenditure of energy.

The process disadvantages are as follows:

a. **Cost.** Both capital, and operation and maintenance costs, including costs for supplemental fuel, are generally higher than for other disposal alternatives.
b. **Operating problems.** High temperature operations create high maintenance requirements and can reduce equipment reliability.
c. **Staffing.** Highly skilled and experienced operators are required for high temperature processes. Municipal salaries and operator status may have to be raised in many locations to attract the proper personnel.
d. **Environmental impacts.** Discharges to the atmosphere (particulates and other toxic or noxious emissions), surface water (scrubbing water), and land (furnace residues) may require extensive treatment to assure protection of the environment (3).

This chapter describes both proven high temperature processes, and those having a high probability of success, as indicated by current research. Multiple hearth and fluid bed furnaces, the most commonly used sludge combustion equipment in the United States, Europe, and Great Britain, are discussed. New thermal processes for wastewater solids reduction are also described. These processes include starved-air combustion and co-combustion of sludges and other residues. The chapter also presents examples that illustrate the methodology used in selecting and designing processes and equipment.

2. PRINCIPLES OF HIGH TEMPERATURE OPERATIONS—COMBUSTION FACTORS

Combustion is the rapid exothermic oxidation of combustible elements in fuel. Incineration is complete combustion. Classical pyrolysis is the destructive distillation, reduction, or thermal cracking and condensation of organic matter under heat and/or pressure in the absence of oxygen. Partial pyrolysis, or starved-air combustion, is incomplete combustion and occurs, when insufficient oxygen is provided to satisfy the combustion requirements. The basic elements of each process are shown in Fig. 1. Combustion of wastewater solids, a two-step process, involves drying followed by burning. For further useful information *see* ref. 4.

2.1. Sludge Fuel Values

A value commonly used in sludge incineration calculations is 10,000 Btu/lb of combustibles (Table 1). It is important to clearly understand the meaning of combustibles. For combustion processes, solid fuels are analyzed for volatile solids and total combustibles. The difference between the two measurements is the fixed carbon. Volatile solids are determined by heating the fuel in the absence of air. Total combustibles are determined by ignition at 1336°F (725°C). By definition, the difference in weight loss is the fixed carbon. In the volatile solids determination used in sanitary engineering (5), sludge is heated in the presence of air at 1021°F (550°C). This measurement is higher than the volatile solids measurement for fuels, and includes the fixed carbon. Numerically, it is nearly the same as the combustibles measurement. In the following, discussion, if "volatile solids" is

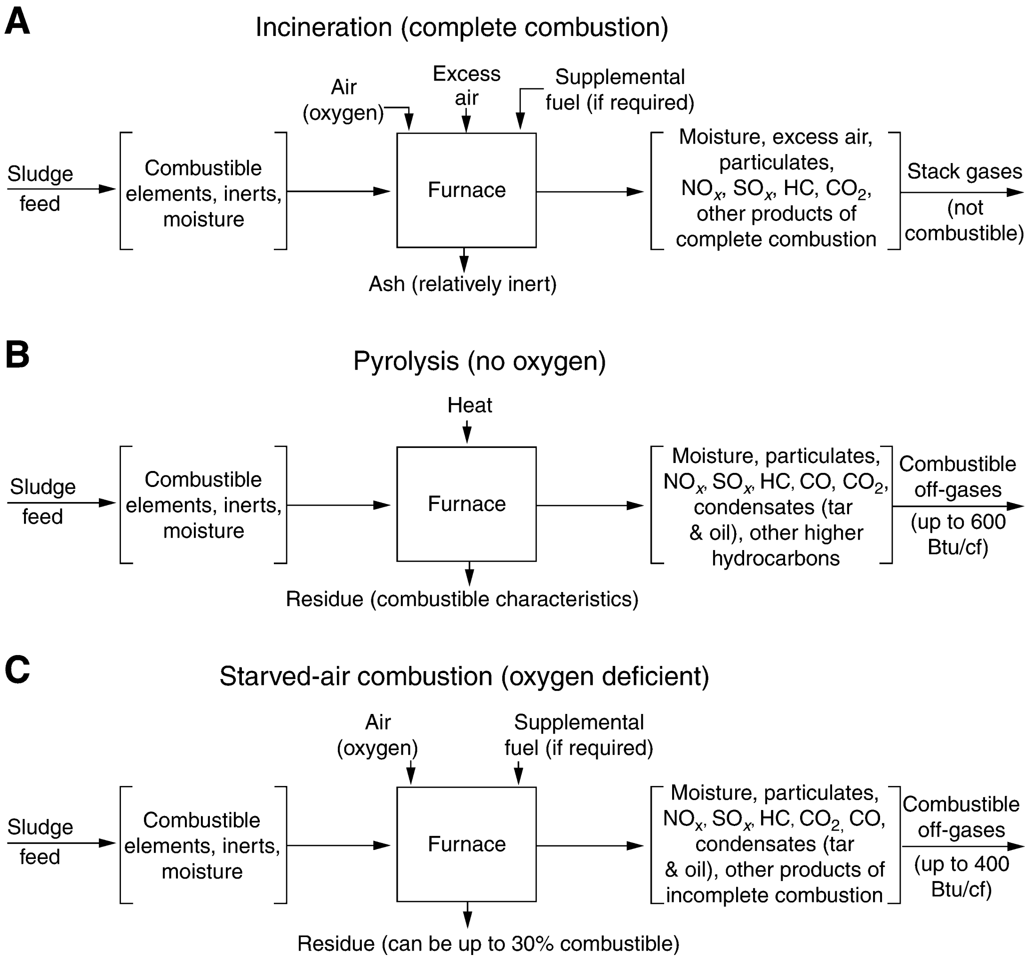

Fig. 1. Basic elements of high temperature processes.

used in the sense of the fuels engineer, it will be followed parenthetically by the designation "fuels usage." If the term "volatile solids" or "volatiles" is used without designation, it will indicate sanitary engineering usage and will be used synonymously with "combustibles."

The amount of heat released from a given sludge is a function of the amount and types of combustible elements present. The primary combustible elements in sludge and in most available supplemental fuels are fixed carbon, hydrogen, and sulfur. Because free sulfur is rarely present in sewage sludge to any significant extent, and because sulfur is being limited in fuels, the contribution of sulfur to the combustion reaction can be neglected in calculations without compromising accuracy. Similarly, the oxidation of metals contributes little to the heat balance and can be ignored.

As shown in Table 2, solids with a high fraction of combustible material; for example, grease and scum, have high fuel values (6). Those which contain a large fraction of inert materials; for example, grit or chemical precipitates, have low fuel values. Chemical precipitates may also exert appreciable heat demands when undergoing high temperature decomposition. This further reduces their effective fuel value.

Table 1
Chemical Reactions Occurring During Combustion

Reaction	High heat value of reaction[a,b]
$C + O_2 \rightarrow CO_2$	–14,100 Btu/lb of C
$C + 1/2O_2 \rightarrow CO$	–4000 Btu/lb of C
$CO + 1/2O_2 \rightarrow CO_2$	–4400 Btu/lb of CO
$H_2 + 1/2O_2 \rightarrow H_2O$	–61,100 Btu/lb of H_2
$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	–23,900 Btu/lb of CH_4
$2H_2S + 3O_2 \rightarrow 2SO_2 + 2H_2O$	–7100 Btu/lb of H_2S
$C + H_2O$ (gas) $\rightarrow CO + H_2$	+4700 Btu/lb of C
Sludge combustibles $\rightarrow CO_2 + H_2O$	–10,000 Btu/lb of combustibles

[a] Negative sign convention indicates an exothermic reaction.

[b] High heat value assumes the latent heat of water generated is available for use: conversely, low heat values assumes the latent heat of water is not available hence no water is condensed.

1 Btu/lb = 2324 J/kg.

Table 2
Representative Heating Values of Some Sludges

Material	Combustibles (%)	High heating value (Btu/lb of dry solids)
Grease and scum	88	16,700
Raw wastewater solids	74	10,300
Fine screenings	86	9000
Ground garbage	85	8200
Digested sludge	60	5300
Chemical precipitated solids	57	7500
Grit	33	4000

1 Btu/lb = 2324 MJ/kg (6).

The following are experimental methods from which sludge heating value may be estimated or computed:

a. Ultimate analysis—an analysis to determine the amount of basic feed constituents. These constituents are moisture, oxygen, carbon, hydrogen, sulfur, nitrogen, and ash. In addition, it is typical to determine chloride and other elements that may contribute to air emissions or ash disposal problems. Once the ultimate analysis has been completed, Dulong's formula Eq. (1) can be used to estimate the heating value of the sludge. Dulong's formula is:

$$\frac{\text{Btu}}{\text{lb}} = 14{,}544\ \text{C} + 62{,}208\left(\text{H}_2 - \frac{\text{O}_2}{8}\right) + 4050\ \text{S} \tag{1}$$

where C, H_2, O_2, and S represent the weight fraction of each element determined by ultimate analysis. This formula does not take into account endothermic chemical reactions that occur with chemically conditioned or physical–chemical sludges. The ultimate analysis is used principally for developing the material balance, from which a heat balance can be made.

b. Proximate analysis—a relatively low-cost analysis in which moisture content, volatile combustible matter, fixed carbon, and ash are determined. The fuel value of the sludge is calculated as the weighted average of the fuel values of its individual components.
c. Calorimetry—this is a direct method in which heating value is determined experimentally with a bomb calorimeter. Approximately 1 g of material is burned in a sealed, submerged container. The heat of combustion is determined by noting the temperature rise of the water bath. Several samples must be taken and composited to obtain a representative 1 g sample. Several tests should be run, and the results must be interpreted by an experienced analyst. New bomb calorimeters can use samples up to 25 g and this type of unit should be used where possible.

The above tests give approximate fuel values for sludges, and allow the designer to proceed with calculations which simulate operations of an incinerator. If a unique sludge will be processed, or unusual operating conditions will be used, pilot testing is advised. Many manufacturers have test furnaces especially suited for pilot testing.

2.2. Oxygen Requirements for Complete Combustion

Air is the normal source of oxygen for combustion; although pure oxygen feed systems are sometimes used. Theoretical air and oxygen requirements for the combustion reactions are shown in Table 1. For rigorous analyses, the constants given in Table 3 should be used. For general applications in which fuel oil, methane, and/or sludge are used, a rule of thumb is that they require 7.5 lb (3.4 kg) of air to release 10,000 Btu (10.55 MJ) from sludge or supplemental fuel (7).

In practice, incinerator operations require air in excess of theoretical requirements for complete combustion. Excess air added to the combustion chamber increases the opportunity for contact between the fuel and oxygen. To ensure complete combustion, it is necessary to maintain 50–150% excess air over the stoichiometric amount required in the combustion zone. When the amount of excess air is inadequate, only partial combustion of carbon occurs, and carbon monoxide, soot, and odorous hydrocarbons are produced.

The excess air required for complete combustion adversely affects the cost of operation, because additional heat is needed to raise the excess air temperature to that of the exhaust gases. Supplemental fuel may be needed to furnish this additional heat. Thermal economy therefore demands that excess air be held to the minimum value required to effect complete combustion. The amount of excess air required varies with the type of incineration equipment, the nature of the sludges to be incinerated, and the disposition of the stack gases. The impact of excess air use on supplemental fuel requirements in sludge incineration is shown in Fig. 2.

2.3. Factors Affecting the Heat Balance

The heat released by burning the wastewater solids must be sufficient to raise the temperatures of all substances entering from ambient levels to those of the exhaust and

Table 3
Theoretical Air and Oxygen Requirements for Complete Combustion

	lb/lb of substance	
Substance	Air	Oxygen
Carbon	11.53	2.66
Carbon monoxide	2.47	0.57
Hydrogen	34.34	7.94
Sulfur	4.29	1
Hydrogen sulfide	6.10	1.41
Methane	17.27	3.99
Ethane	16.12	3.73
Ammonia	6.10	1.41

1 lb/lb = 1 kg/kg.

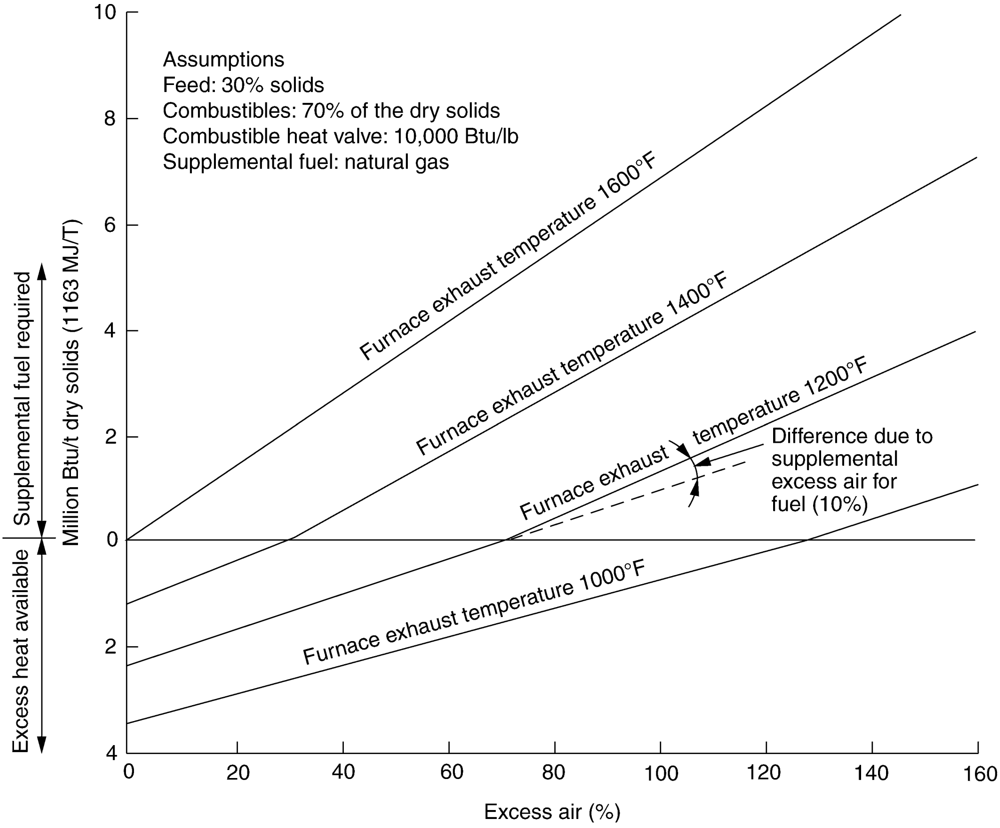

Fig. 2. Effect of excess air and excess temperature on supplemental fuel requirements.

solid residue streams. Also, any radiant heat loss from the combustion structure must be included. If the heat is sufficient, the process is termed autogenous. If it is not sufficient, supplemental fuel must be burned to make up for the heat deficit.

A number of variables influence the amount of supplemental fuel required. As shown in Fig. 2, the amount of excess air required to produce complete combustion has an important effect. Water associated with the sludge also exerts significant demands. For example, it takes almost 2000 Btu/lb (4.64 MJ/kg) to vaporize water and raise the temperature of the water vapor to exhaust temperatures. When allowances are made for radiation losses and for heating of gas streams and sludge feed solids, it is found that approx 3500 Btu (3.69 MJ) are required for every pound (0.45 kg) of water evaporated in a multiple hearth furnace (8).

The following example illustrates how the feed solids concentration required for autogenous combustion is determined.

2.4. Example

A designer uses a proximate analysis to derive the following values for a given sludge volatile solids content:

a. Fuels usage = 66%.
b. Fixed carbon content = 11%.
c. Inert content = 23%.

The sludge is to be dewatered and burned in a multiple hearth incinerator. Determine the solids concentration required for autogenous combustion in a multiple hearth incinerator. The sludge heating value can be estimated by multiplying the approximate fuel value of sludge, 10,000 Btu/lb (23.2 MJ/kg), by the combustible fraction in the sludge. In this example, the combustible fraction is the sum of the volatile solids (fuels usage) and fixed carbon, or 77%. Therefore, sludge heating value is: 10,000 Btu/lb × 0.77 = 7700 Btu/lb (17.89 MJ/kg).

The minimum percent sludge solids required to maintain autogenous combustion can be determined by equating the heat released by combustion to the heat required by the water. Therefore:

$$P \times Q = (100 - P)W \tag{2}$$

where P is the minimum percent dry solids in sludge required for autogenous combustion (%); and Q is the fuel value of sludge (Btu/lb of dry solids); and W is the heat required to evaporate 1 lb of water in a multiple hearth furnace (Btu).

Equation 2 above is solved for P:

$$P = \frac{W}{Q + W} \times 100 \tag{3}$$

For this example:

$$P = \frac{3500}{7700 + 3500} \times 100 = 31.3\%$$

If the solids could be dewatered to 31.3%, they could be combusted autogenously. However, feed solids concentrations of this magnitude are seldom achieved without chemical conditioning. Allowances for the effect of chemical conditioning should therefore be made. Assume conditioning requirements are 25% lime and 3% ferric chloride by weight of dry solids fed. Therefore, for every 100 lb (45.4 kg) of sludge dewatered, 28 lb (12.7 kg) of chemicals are added. Assuming there is no heating value in the lime and ferric chloride, the combustible fraction of the feed solids is reduced to 100/128 × 0.77 = 60%, and the sludge heating value is 6000 Btu/lb (13.9 MJ/kg). Using Eq. (3), the dewatered sludge must be 36.8% solids to be autogenous.

Figure 3 shows a family of curves that can be used to calculate the minimum percent of solids required at various dry solids heating values. This method of estimating takes into account the effect of moisture content, inerts, and combustibles on the combustion process and can be used for basic sizing prior to detailed analysis.

For example, in the above analysis, a sludge heating value of 6000 Btu/lb of solids (13.9 MJ/kg) was calculated. From Fig. 3, the 6000 Btu/lb (13.9 MJ/kg) curve crosses the breakeven point at approx 36% dry solids. The importance of dewatering the sludge is illustrated in Fig. 4. The amount of supplemental fuel required is plotted as a function of feed moisture content and combustible solids concentration.

The amount of supplemental fuel can be reduced, if heat can be recovered from the process exhaust gases and reused. As an example, heat may be transferred from the furnace flue gas to incoming combustion air by means of heat exchangers (recuperateors). Although energy recovery can significantly improve thermal efficiency, heat recovery equipment can be expensive and can only be recommended after complete economic evaluation.

3. TECHNOLOGY REVIEW

This section discusses four major technologies for high temperature conversion of sludge are as follows:

a. Incineration of sludge fluidized bed furnace (FBF; 9).
b. Incineration of sludge multiple hearth furnace (MHF; 10).
c. Codisposal by combustion (11).
d. Starved air combustion of sludge (SAC; 12).

3.1. Incineration of Sludge FBF

3.1.1. Description

The first fluidized bed wastewater sludge incinerator was installed in 1962. There are now many units operating in the US with capacities of 200–1000 lb/h of dry solids. Sludge incineration is a two-step process involving drying and combustion after preliminary dewatering. A typical sludge contains 75% water and 75% volatiles in dry solids. Self-sustained combustion without supplementary fuel is often possible with dewatered raw sludges having a solids concentration more than 30%.

The FBF (Fig. 5) is a vertically oriented, cylindrically shaped, refractory lined, steel shell, which contains a sand bed and fluidizing air distributor. The FBF is normally available in diameters of 9–25 (ft) and heights of 20–60 (ft). There is one industrial unit operating with a diameter of 53 ft. The sand bed is approx 2.5 ft thick and rests on a refractory-lined air distribution grid containing tuyeres through which air is injected at a pressure of 3–5 psi to fluidize the bed. Bed expansion is approx 60–100%. The temperature of the bed is controlled between 1400 and 1500°F by auxiliary burners and/or a water spray or heat removal system above the bed. Ash is carried out through the top of the furnace and is removed by air pollution control

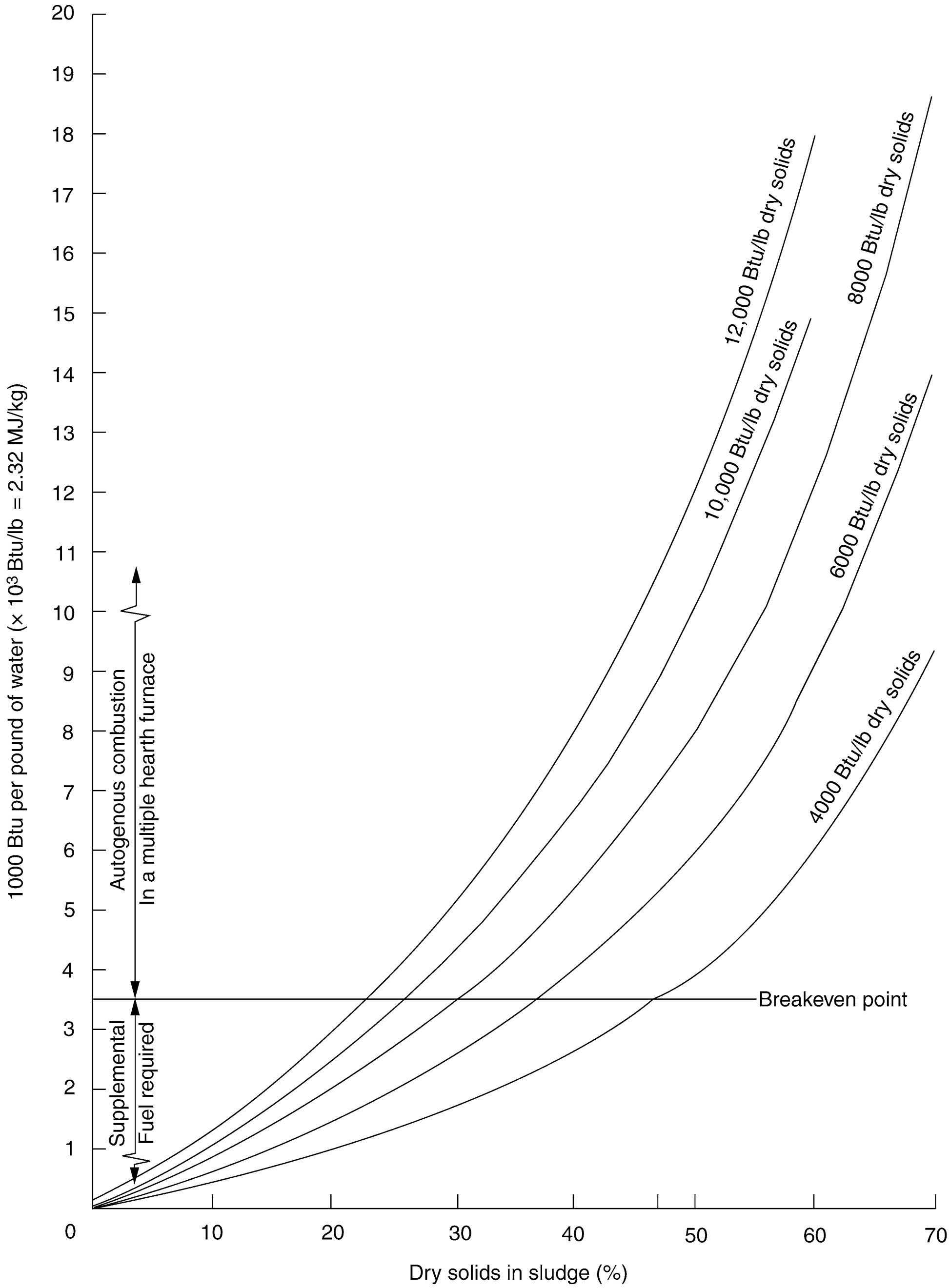

Fig. 3. Effect of dry solids heating value and sludge moisture on capability for autogenous combustion.

devices, usually wet venturi scrubbers. Sand is lost by attrition at a rate of approx 5% of the bed volume for every 300 h of operation. Furnace feed can be introduced either above or directly into the bed depending on the type of feed. Generally, sludge is fed directly into the bed.

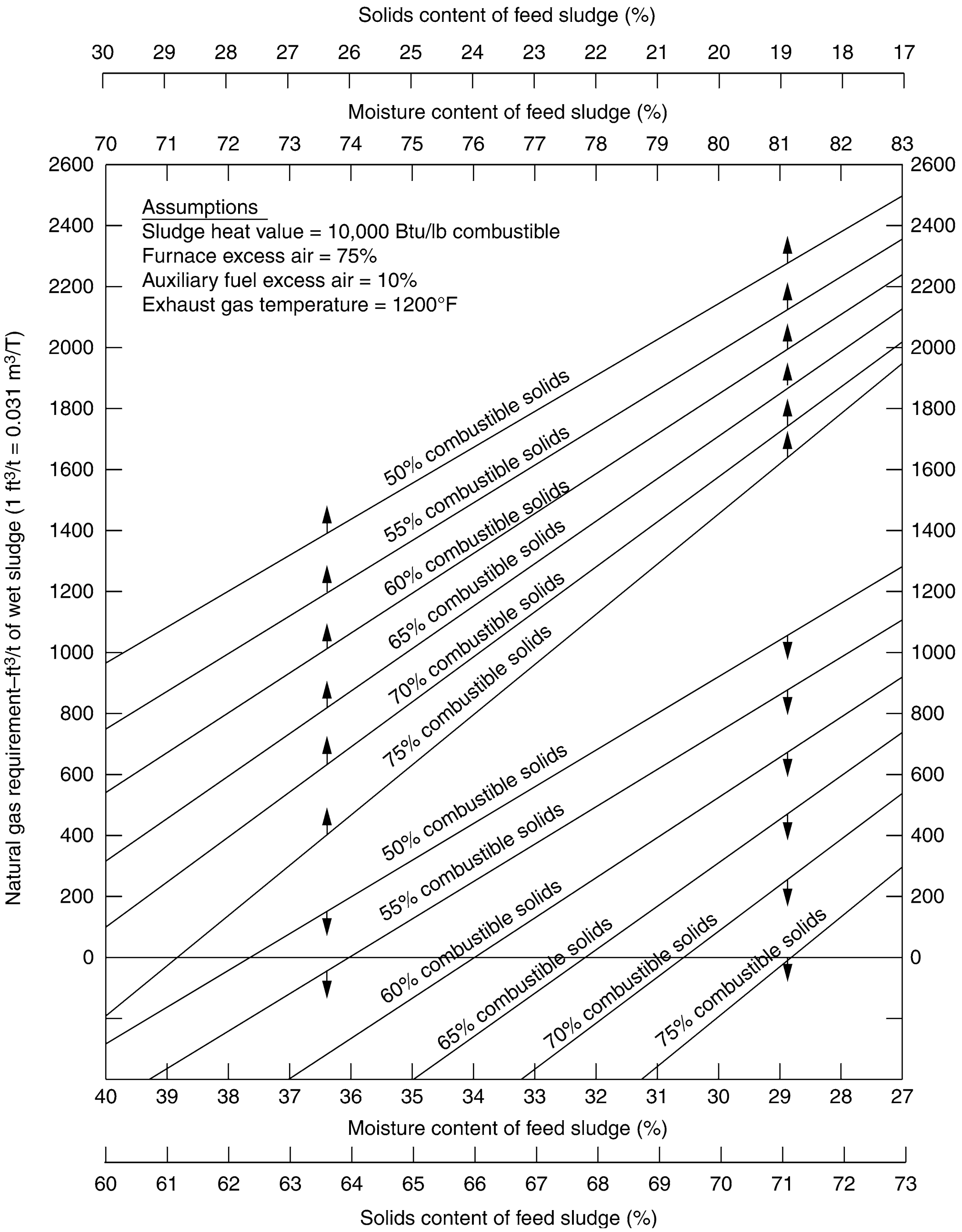

Fig. 4. Effect of sludge moisture content and combustible solids content on supplemental fuel consumption. 1 t = 2000 lb; 1 T = 2204.6 lb = 1000 kg.

Excess air requirements for the FBF vary from 20 to 40%. It requires less supplementary fuel than a MHF. An oxygen analyzer in the stack controls the air flow into the reactor and the auxiliary fuel feed rate is controlled by a bed temperature controller.

Start-up fuel requirements are very low, and no fuel is required for start-up following an overnight shutdown. The FBF is very attractive for intermittent operation. Afterburners

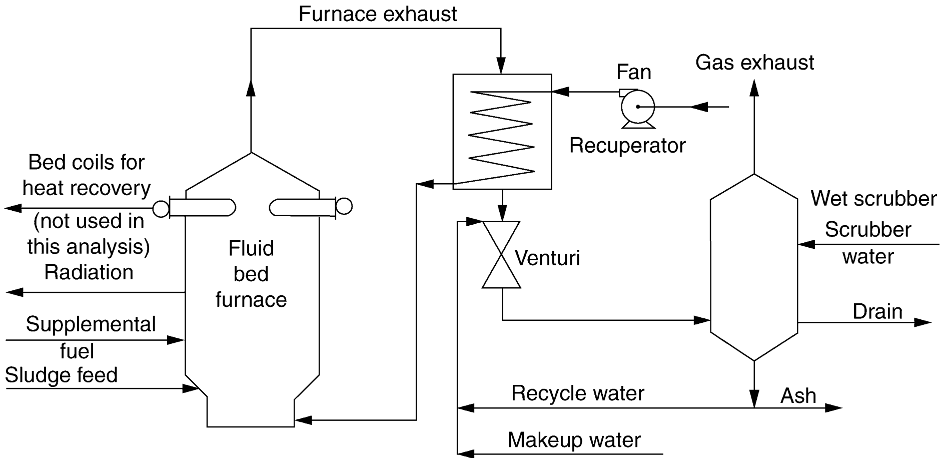

Fig. 5. Incineration of sludge in a fluidized bed furnace.

are not required to comply with air pollution regulations. An air preheater is used in conjunction with a fluidized bed to reduce fuel costs. Also, cooling tubes may be submerged in the bed to provide energy recovery.

3.1.2. Applications

The FBF are most suitable for:

a. Reduction of sludge volume, thereby reducing land requirements for disposal.
b. Energy recovery potential.
c. Where hauling distances to disposal sites are long.
d. Where regulations concerning alternative methods are prohibitive.

3.1.3. Performance

The mass of dry solids is reduced to 25–35% of the amount entering the unit. Because a minimum amount of air is always required for bed fluidization, fan energy savings during load turndown (i.e., sludge feed reduction) are minor. It is generally not cost effective for small plants.

Some extensive maintenance problems have occurred with air preheaters. Scaling of the venturi scrubbers has also been a problem. Screw feeds and screw pump feeds are both subject to jamming from either overdrying of the sludge feed at the incinerator, or from silt carried into the feed system with the sludge. Another frequent problem has been the burnout of spray nozzles or thermocouples in the bed.

3.1.4. Design Criteria

The design criteria are as follows:

a. Bed loading rate (50–60 lb wet solids/ft^2/h).
b. Superficial bed velocity (0.4–0.6 ft/s).
c. Sand effective size (0.2–0.3 mm) (uniformity coefficient 1.8).
d. Operating temperature (1400–1500°F normally [2200°F maximum]).
e. Bed expansion (80–100%).
f. Sand loss (5% of bed volume per 300 h of operation).

Table 4
Fluidized Bed Furnace Operations Data

Plant flow (MGD)	Operating (d/wk)	Operating (h/d)
0.1	1	20
1	7	20
10	7	20
100	7	20

3.1.5. Environmental Impact

Particulate collection efficiencies of 96–97% are required to meet current emission standards. There is little data on the amount of toxic metals which are volatilized and discharged. Limited test data (8) indicate that 4–35% of the mercury entering an incinerator with emission controls will volatilize and be emitted to the atmosphere (excluding particulate forms). Appreciable gaseous emissions of CO, HCl, $SO_{2,}$ and NO_2 may require additional air pollution control measures. Pesticides and polychlorinated biphenyls (PCBs) may also be found in the sludge, but tests indicate that they can be destroyed during incineration and should not be a problem.

3.1.6. Energy Requirements

Using the design basis below, electrical energy requirements are approx 90,000 kWh/yr/dry t/d or 85,000 kWh/yr/MGD plant flow. Fuel requirements are approx 90 gal/dry t of sludge or 13×10^6 Btu/dry t. Fuel requirements are very sensitive to the moisture content of the sludge and other factors. As a result, adjustments should only be made after detailed study for each case.

3.1.7. Design Basis for Costs

Construction costs include those for the reactor, air blowers, and accessories, preheaters, scrubbers, fuel pumps, and building. Costs are calculated for undigested dewatered primary and secondary sludge (1900 lb/MGD at 20% solids; 75% volatile). Table 4 shows FBF operations data.

3.2. Incineration of Sludge MHF

3.2.1. Description

Sludge incineration is a two-step process involving drying and combustion after preliminary dewatering. A typical sludge is 80% water and has a dry solids volatility of 75%. Self-sustained combustion without supplementary fuel is often possible with dewatered raw primary sludges which can frequently be dewatered to 30% solids.

The MHF (Fig. 6) is a vertically oriented, cylindrically shaped, refractory lined, steel shell (diameter = 4–25 ft) containing 4–13 horizontal hearths positioned one above the other. The hearths are constructed of high-heat-duty fire bricks and in special fire brick shape. Sludge is raked radially across the hearths by rabble arms which are supported by a central rotating shaft that runs the height of the furnace. The cast iron shaft is motor driven with provision for speed adjustment from 0.5 to 1.5 rpm. Sludge is fed to the top hearth and proceeds downward through the furnace from hearth to hearth. Inflow hearths have a central port through which sludge passes to the next lower hearth.

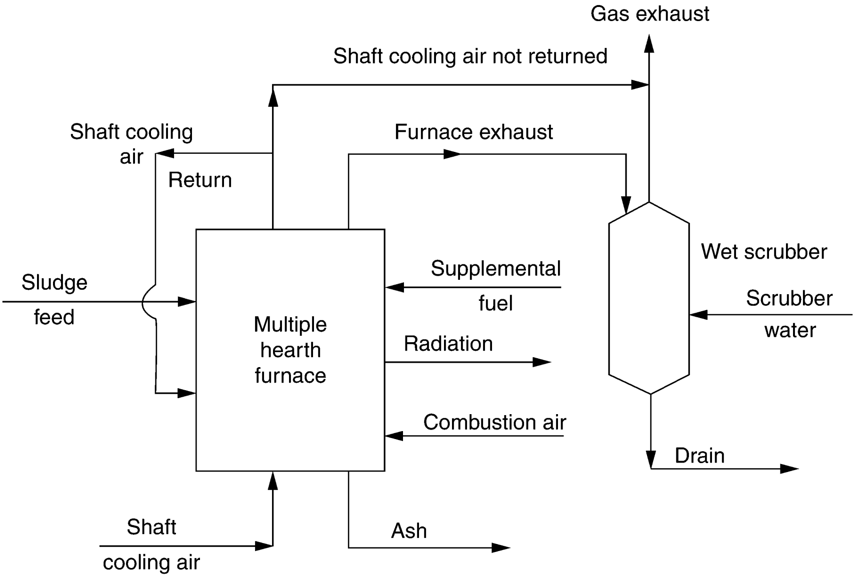

Fig. 6. Incineration of sludge in a multiple hearth furnace.

Outflow hearths have ports on their periphery. These ports tend to regulate gas velocities also. The central shaft contains internal concentric flow passages through which air is routed to cool the shaft and rabble arms. The flow of combustion air is countercurrent to that of the sludge. Gas or oil burners are provided on some hearths for start-up and/or supplemental use as required.

The rabble arms provide mixing action as well as movement to the sludge so that a maximum sludge surface is exposed to the hot furnace gases. Because of the irregular surface left by the rabbling action, the surface area of sludge exposed to the hot gases is as much as 130% of the hearth area. While there is significant solids–gas contact time on the hearths, the overall contact time is actually still greater, owing to the fall of the sludge from hearth to hearth through the countercurrent flow of hot gases.

The various phases of the incineration process occur in three zones of the MHF. The drying zone consists of the upper hearths, the combustion zone consists of the central hearths, and the cooling zone consists of the lower hearths. Temperatures in each zone are:

a. Drying zone—sludge about 100°F; air about 800°F.
b. Burning zone—sludge and air about 1500°F.
c. Cooling zone—sludge about 400°F; air about 350°F.

To meet air emission restrictions, off-gases are treated with catalytic combustion, waste-gas flares, and direct flame incinerators. Waste gas flares are used for nonhazardous waster with a high organic content. Direct flame incinerators operating at 1000–1500°F are used for waste gasses containing particles (13).

3.2.2. *Applications*

The MHF is a widely used form of wastewater sludge incinerator in the United States. New developments in the technology include the use of oxygen-enhanced incineration. A study performed by Praxair demonstrated improved performance, including (14):

a. Increase long-term throughput.
b. Reduced emissions.
c. Reduced natural gas consumption.
d. Greater flexibility and more rapid response to changing input stream conditions.

The MHF is mostly used for:

a. Reduction of sludge volume, thereby reducing land requirements for disposal.
b. Energy recovery potential.
c. Plants that have long hauling distances to land or ocean disposal sites.
d. Where regulations prohibit these alternate disposal methods.

3.2.3. Performance

Dry solids are reduced to 20–25% of the mass entering the unit. In the case of the Fairfax County, Virginia plant, the final product is an inert ash having 10% of the original sludge volume (15). The recoverable heat ranges from 18% of the total heat input (sludge and supplementary fuel) at 20% solids concentration to 45% of the total heat input at 40% solids concentration.

Capacities of MHFs vary from 200 to 8000 lb/h of dry sludge. Maximum operating temperatures are limited to 1700°F. With high-energy feeds there may be operational problems. The MHF requires 24–30 h for furnace warm-up or cool-down to avoid refractory problems. Failure of rabble arms and hearths has also been encountered. Nuisance shutdowns have also occurred owing to ultraviolet flame scanner malfunctions. Thickening and dewatering pretreatment is required.

3.2.4. Design Criteria

a. Maximum operating temperature (1700°F).
b. Hearth loading rate (6–10 lb wet solids/ft^2/h with a dry solids concentration of 20–40%).
c. Combustion air flow (12–13 lb/lb dry solids).
d. Shaft cooling air flow (1/3–1/2 of combustion air flow).
e. Excess air (75–100% [16]).

3.2.5. Environmental Impact

Particulate collection efficiencies of 96–97% are required to meet current emission standards. There is little data on the amount of toxic metals which are volatilized and discharged. Limited test data (8) indicate that 4–35% of the mercury entering an incinerator with emission controls will volatilize and be emitted to the atmosphere (excluding particulate forms). Gaseous emissions of CO, HCl, $SO_{2,}$ and NO_2 are expected to be acceptable. Pesticides and PCBs may be found in the sludge, but tests indicate that they can be destroyed during incineration and should not be a problem.

3.2.6. Energy Requirements

Electrical energy requirements are approx 31,000, 135,000, and 1,250,000 kWh/yr for l, 10, and 100 MGD plant flows respectively. Fuel requirements for startup and incineration amount to approx 4500×10^6 Btu/yr/MGD. Fuel requirements are very sensitive to the moisture content of the sludge and other factors. As a result, adjustments should only be made after detailed study for each case.

Table 5
Multiple Hearth Furnace Operations Data

Plant flow (MGD)	Operating (d/wk)	Operating (h/d)
0.1	1	20
1	7	20
10	7	20
100	7	20

3.2.7. Design Basis for Costs

Construction costs commonly include those from incinerator, building, sludge conveyor, ash handling equipment, and gas scrubbers. Costs are usually calculated for undigested dewatered primary and secondary sludge (1900 lb/MGD at 20% solids, 75% volatile). Table 5 shows MHF operation data.

3.3. Codisposal by Combustion

3.3.1. Description

Codisposal of sludge by starved air combustion (SAC) is an extension of a process using waste materials such as municipal solid waste, wood waste, farm waste, and so on. as fuel additives (Fig. 7). This allows operation of the unit without auxiliary fossil fuel in the case of high moisture content sludge or sludges with low solids heating value.

At a test run at the Central Contra Costa Sanitary District operated wastewater treatment plant in Concord, California, a 16-ft-diameter, 6-hearth MHF processed a combination of sludge and refuse derived fuel (RDF). Mixed municipal refuse was shredded, classified, and screened prior to addition to the MHF where the RDF was the light fraction from the air classifier. The sludge had a solids content of 16%, a volatile solids content of 75%, and a heating value of 9000 Btu/lb dry solids, whereas the RDF had a solids content of 75%, very few inerts, and a heating value of 7500 Btu/t of dry solids. The furnace feed rate varied from pure sludge to pure RDF. A combustible gas was produced with a heating value of 130 Btu/sdft^3. This combustible gas could be fired in a waste heat boiler for steam production, used as the fuel for a lime recalcination furnace, or used for space heating.

During the test, the RDF could be fed to hearths 3 or 1. Sludge was always fed to hearth 1. Temperatures were maintained by controlling the amount of air fed to the furnace. The off-gases from the furnace were allowed to burn in an afterburner with the introduction of combustion air. Afterburner temperatures were approx 2200°F, although the gas could be combusted to produce a temperature as high as 2500°F with no supplemental fuel addition.

The major shaft furnace systems available, the Purox (oxygen enrichment) and Torrax (regenerative heat recovery) units by Union Carbide Co. and The Carborundum Co., respectively, have similar basic operating principles. Refuse is charged at the top of the refractory lined shaft, providing a seal, and, as it descends through the furnace, hot pyrolytic gases from the slagging and combustion zones move in a countercurrent direction, thus providing preignition and drying of the sludge and refuse. Preheated air or oxygen-enriched air is injected into the combustion zone at the base of the shaft furnace, where combustion of the pyrolyzed char occurs.

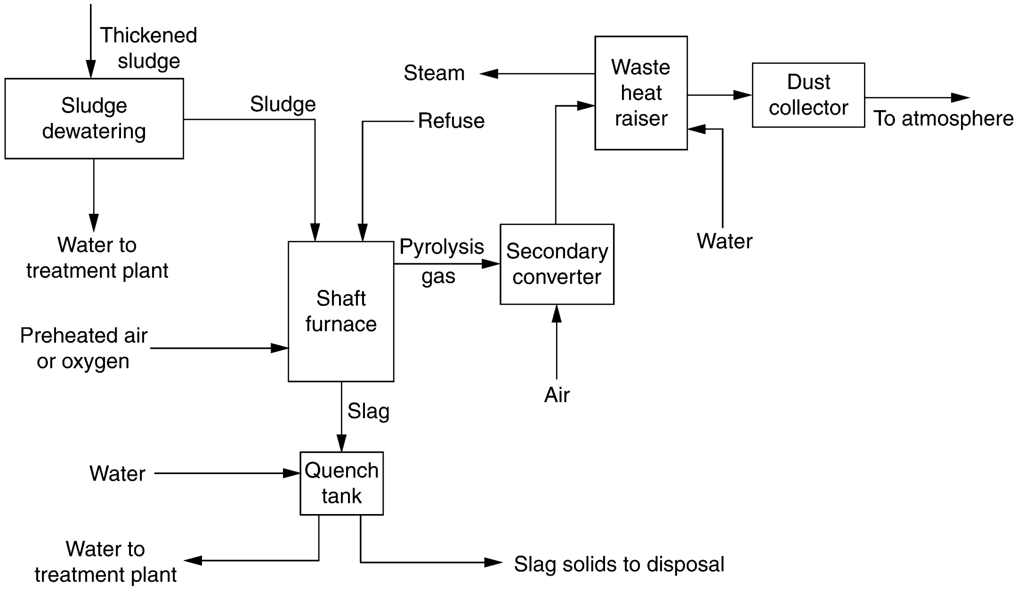

Fig. 7. Codisposal by starved air combustion.

Preheating and oxygen enrichment serve essentially the same purpose—maintaining a furnace temperature high enough (2500–3000°F) to form a slag and to produce a pyrolysis gas with as high a heating value as possible. The slag formed is virtually free of combustibles. The cooled gases (low heating value), after preliminary cleaning, can be burned in a secondary combustion chamber with energy recovery in the form of a waste-heat boiler.

3.3.2. *Applications*

The major applications are as follows:

a. Reduction in volume of two solid waste streams.
b. Energy conservation.
c. Use a single device to dispose of two waste products, thereby.
d. Realizing capital and operating cost benefits.

3.3.3. *Performance*

Quantity of dry solids (combined sludge and RDF) is reduced to 19–26% of amount entering reactor (16). Institutional constraints may hamper implementation; for instance, in many localities, wastewater treatment and solid waste disposal are controlled by different governmental agencies. Many communities have long-term (15–20 yr) contracts with private firms for refuse handling and disposal which define ownership of the refuse. In shaft furnaces, proper temperatures in the slag tap area must be maintained to prevent slag freezing. The Purox system and the MHF require a shredded refuse feed.

3.3.4. *Design Parameters and Reliability*

For MHF, the hearth sludge loading rate is 11–13 lb wet solids/h ft^2 (16). MHF units have experienced failures of rabble arms and hearths along with nuisance trips owing to flame scanner malfunction. Fluid bed units have experienced scaling of scrubbers with

Table 6
Capital Cost for Codisposal by Starved Air Combustion

Item	Cost (USD)
Two shaft furnaces[a]	44,750,000
Additional building[a]	3,536,000
Direct construction cost (DCC)	48,286,000
Design, construction management	
Start-up (15% DCC)	7,243,000
Land (65,000/acre USD)[b]	250,000
Legal fees (3% DCC)	1,449,000

[a] Values adjusted form original mid-1975 (based on ENR 2212) to August 2003 (based on *Chemical Engineering* Plant Cost Index); (31,32).

[b] Greater Syracuse Economic Growth Council (33).

bed media and plugging of sludge feed systems. Freezing of slag bed and contamination of slag bed with refuse have been experienced on shaft furnace systems.

3.3.5. Environmental Impact

The impact will depend on compositions of feed material and operating conditions. Hydrocarbons and CO emissions are not of concern since product fuel gas is sought. SO_2 and NO_x emissions from the reactor may be reduced relative to coincineration owing to deficiency of oxygen. However, higher temperatures in the afterburner could cause a NO_x problem.

3.3.6. Energy Requirements

Electrical power requirements are 51 kWh/t of refuse and sludge. Auxiliary fuel requirements are 6.2 gal of fuel oil or 780,000 Btu/t of refuse and sludge.

3.3.7. Design basis for Costs

The capital and operating costs for codisposal by SAC are shown in Tables 6 and 7 respectively. Capital costs include: equipment, labor and materials for installation, construction overhead and contingency at 15% of each equipment module for

a. Coincineration/SAC (sludge and refuse).
b. Design capacity: 600 t/d refuse.
c. 224 t/d sludge.
d. Annual capacity 206,000 t.

Manpower costs include: four shifts (7 d/wk operation), supervision, and maintenance. About 20% is added for overtime, vacations, holidays, and so on.

3.4. SAC of Sludge

3.4.1. Description

The process utilizes equipment and process flows similar to incineration except that less than the theoretical amounts of air for complete combustion is supplied (Fig. 8). Autogenous SAC can be achieved with a sludge solids concentration more than 25%. For lower concentrations, an auxiliary fuel may be required, depending on the percent volatiles in the solids. High temperatures decompose or vaporize the solid components of this

Table 7
Operating Cost for Codisposal by Starved Air Combustion

Item	Cost/t[a] (USD)	Total annual cost (USD)
Manpower (42 employees)[b]	5.90	1,214,000
Power (1450 kWh/h)[c]	3.40	699,800
Water/sewer (300 gal/min)[d]	0.64	133,600
Auxiliary fuel and heating (1057, 400 gal/yr)[e]	9.49	1,618,000
Maintenance (2.5% DCC)	5.82	1,207,000
Overhead (1% DCC)	2.33	482,900
Residue (488 t/d)[f]	1.58	325,700
Total operating cost	29.16	5,681,000

[a] Based on Annual Throughput.

[b] Values adjusted form original mid-1975 (based on ENR 2212) to August 2003 (based on *Chemical Engineering* Plant Cost Index; 31,32).

[c] Based on (0.067 USD/kWh) Electric Rate 6, Nicor, Inc. (34).

[d] Based on (1.03 USD/1000 gal) Greater Syracuse Economic Growth Council (33).

[e] Based on (6.70 USD/MMBtu for natural gas) Nicor, Inc. (34).

[f] Values adjusted form original mid-1975 (based on ENR 2212) to August 2003 (based on *Chemical Engineering* Plant Cost Index). Values will vary significantly depending on toxicity of residue (31,32).

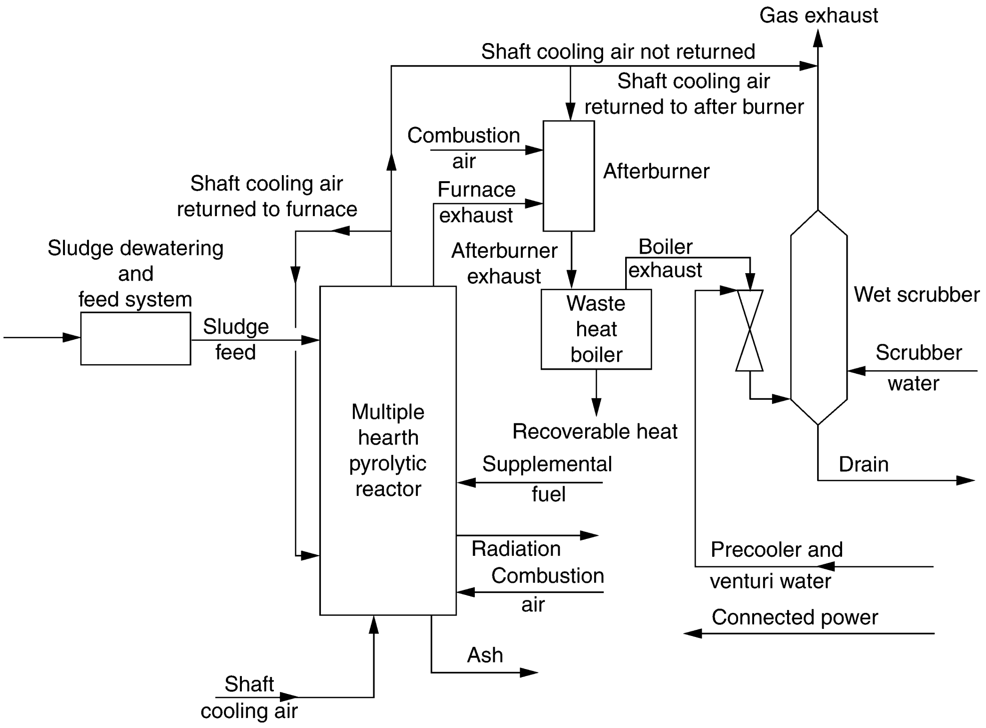

Fig. 8. Starved air combustion of sludge.

sludge. The gas phase reactions are pyrolytic or oxidative, depending on the concentration of oxygen remaining in the stream. Under proper control, the gas leaving the vessel is a low Btu fuel gas that can be burned in an afterburner to produce power and/or thermal energy.

Some processes utilize pure oxygen instead of air and thus produce a higher Btu fuel gas. The solid residue is a char with more or less residual carbon, depending on how much combustion air had to be supplied to reach the proper operating temperature. Because the process is neither purely pyrolytic nor purely oxidative, it is called starved-air combustion or thermal gasification, rather than pyrolysis. Other processes still in the development stage use indirect heating, rather than the partial combustion. These are true pyrolysis processes. SAC reduces the sludge volumes and sterilizes the end product. Unlike incineration, it offers the potential advantages of producing useful byproducts and of reducing the volume of sludge without a large amount of supplementary fuels. The gas which is produced has a heat value up to 130 Btu/standard dry ft^3 using air for combustion, and is suitable for use in local applications, such as combustion in an afterburner or boiler or for fuel in another furnace. SAC has a higher thermal efficiency than incineration owing to the lower quantity of air required for the process. In addition, capital economies can be realized owing to the smaller gas handling requirements.

Furnaces may be operated in one of three modes resulting in substantially different heat generation and residue characteristics. The low temperature char mode only pyrolyzes the volatile material, thereby producing a charcoal-like residue with high ash content. The high temperature char mode produces a charcoal-like material converted to fixed carbon and ash, and the char burned mode reacts away all carbon and produces ash as a residue. Heat recovered is at maximum for the char burned mode, less for the high temperature char mode, and substantially less for the low temperature char mode of operation.

SAC operation has shown the following advantages in addition to which are discussed above: easier to control than a standard incinerator; more stable operation with little response to changes in feed; more feed capacity as compared to an equal area for incineration; all equipments used are currently being manufactured; less air pollutants and easier air pollution control management, lower sludge solids content required for autogenous operation.

3.4.2. Applications and Limitations

The main application is for reduction of sludge volume and the production of fuel gas for a nearby combustor or furnace. Most existing MHFs can easily be retrofitted to operate in the SAC mode.

However there are some significant disadvantages are as follow:

a. Need for afterburner may limit use in existing installations owing to space problems. Relatively large amount of instrumentation is required.
b. Must be very careful of bypass stack exhaust, because furnace exhaust is high in hydrocarbons and may be combustible in air. This may result in bypassing only after afterburning with appropriate emergency control in some areas.
c. Corrosivity of furnace exhaust gases.
d. Combustibles in ash may create ultimate disposal problems.
e. Sludge volume reduction lower than incineration.
f. Requires recovery of the energy in the product gas to fully realize the improved efficiency.

3.4.3. Performance

Unit can operate without auxiliary fuel, including afterburner, with sludge dewatered to the range of 29–39% solids. Based on a limited number of pilot scale tests, the off-gas from an MHF unit operating in the SAC mode, with sludge alone, ranges from 18 to 73 Btu/std ft^3.

Mechanical function of MHF units under the SAC mode is expected to be similar to the conventional operating modes. Increased operating stability is expected to result in higher process reliability. Air pollution can be expected to be less of a problem owing to the lower air flows and the potential for particulate carryover. Data to date indicate conventional equipment can achieve acceptable controls. Depending upon the mode of operation, heavy metals in the sludge can be retained in the residue.

3.4.4. Design Criteria

a. MHF hearth loadings of 9–15 lb wet (22%) solids/ft^2/h.
b. For autogenous combustion, sludge solids content 25–39% depending upon volatility.
c. Off-gas heating value is dependent upon operating mode.

3.4.5. Energy Requirements

a. Electrical energy requirements (23 kWh/t).
b. Auxiliary or startup fuel requirements (3.1 gal fuel oil or 0.43×10^6 Btu/t of sludge).

3.4.6. Design Basis for Costs

a. About 324 t/d design capacity at 40% dry solids.
b. Annual throughput 80,000 t.
c. Direct construction cost (DCC) includes MHF installed, with drives, fans, motor control, gas scrubber, external afterburner, ash handling system, auxiliary fuel system, instrumentation, piping, painting, initial operation and test.

The capital and operating costs for SAC of sludge are shown in Tables 8 and 9 respectively.

4. INCINERATION DESIGN EXAMPLE

To evaluate combustion processes, a designer must determine if the sludge will burn autogenously. He must also assess the effects of different excess air rates, the effects of different types and quantities of supplemental fuel, and combustion air requirements. Useful information on incineration technologies is provided in ref. 17. Very current information is available through the FRTR Remediation Technologies Screening Matrix and Reference Guide 4.0 (18).

Approximate and theoretical methods (TM) for calculating combustion requirements are presented in the following examples. A summary is then provided that compares the results of each method. Either method provides the information necessary for preliminary evaluation and conceptual design of a sludge incinerator. When an ultimate analysis of the sludge is available or a good estimate of sludge constituents can be made, a theoretical analysis is preferred. Useful information on design is available in refs. 19 and 20.

4.1. Problem Statement

The dewatered sludge production rate expected for a wastewater treatment plant is 14,000 lb/h (6350 kg/h) at 20% solids content. The dewatered material is a mixture of undigested primary and wasted activated sludges with a volatile (combustible) content of 77%. The sludge temperature is 60°F (16°C). To limit hydrocarbon emissions, an afterburner is used to heat furnace exhaust gases to 1400°F (760°C). The design is to be based on 100% excess air (two times the theoretical requirement). If supplemental fuel is required, no. 2 fuel oil will be used. 25% excess air will be used for combustion of

Table 8
Capital Cost for Starved Air Combustion of Sludge

Item	Cost (USD)
DCC	4,231,000
Design, construction management (20% DCC)	846,000
Land (65,000 USD/acre)	325,000
Legal fee (3% DCC)	126,900
Bond discount (3% total cost)	165,900
Total cost	5,694,800

[a] Values adjusted form original August 1978 (based on ENR 2829) to August 2003 (based on *Chemical Engineering* Plant Cost Index); (32,35).

[b] Syracuse Economic Growth Council (33).

Table 9
Operating Cost for Starved Air Combustion of Sludge

Item	Cost/t[a] USD	Annual cost (USD)
Manpower (20 employees)	7.33	587,300
Power (210 kWh/h)	1.14	91,320
Water/sewer @ 385 gal/min[d]	2.48	197,100
Auxiliary fuel (250,000 gal/yr)	4.77	382,500
Maintenance (2.5% DCC)	1.32	105,600
Overhead (1% DCC)	0.53	42,240
Residual disposal	1.71	136,500
Total cost	19.28	744,450

[a] Based on 80,000 t/yr throughput.

[b] Values adjusted form original August 1978 (based on ENR 2829) to August 2003 (based on *Chemical Engineering* Plant Cost Index); (32,35).

[c] Based on (0.067 USD/kWh) Electric Rate 6, Nicor, Inc. (34).

[d] Based on (1.03 USD/1000 gal) Greater Syracuse Economic Growth Council (33).

[e] Based on (6.70 USD/MMBtu for natural gas) Nicor, Inc. (34).

[f] Values adjusted form original August 1978 (based on ENR 2829) to August 2003 (based on *Chemical Engineering* Plant Cost Index). Values will vary significantly depending on toxicity of residue (32,35).

the fuel oil. The air temperature is 60°F (16°C); the absolute humidity of air is 0.013 lb of water/lb of dry air. Heat capacities of dry air, water vapor, dry sludge solids, and water are 0.256, 0.50, 0.25, and 1 Btu/lb-°F, respectively, (1.07, 2.1, 1, and 4.2 kJ/kg-°C). The latent heat of water is 970.3 Btu/lb (2253 kJ/kg).

4.2. Approximate Calculation Method

Assuming 10,000 Btu/lb (23.2 MJ/kg) of sludge, the heat content of the sludge is:

$$10{,}000\ \frac{\text{Btu}}{\text{lb}} \times 0.77 = 7700\ \frac{\text{Btu}}{\text{lb}}\ (17.9\ \text{MJ/kg})$$

From Fig. 3, a value of approx 32% solids in the dewatered sludge is required for autogenous combustion. Therefore, supplemental fuel is required and its quantity must be determined. The demand for supplemental fuel equals the heat required minus the heat value of the sludge.

a. **Step 1.** Sludge heating value.
The heating value of the sludge

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.2 \text{ lb solids}}{\text{lb sludge}}\right)\left(\frac{0.77 \text{ lb VS}}{\text{lb solids}}\right)\left(\frac{10{,}000 \text{ Btu}}{\text{lb VS}}\right)$$

$$= 21.56 \times 10^6 \text{ Btu/h (22.75 KJ/h)}$$

b. **Step 2.** Combustion air requirements.
Combustion air requirements.

$$= \left(\frac{21.56 \times 10^6 \text{ Btu}}{\text{h}}\right)\left(\frac{7.5 \text{ lb dry air}}{10{,}000 \text{ Btu}}\right)\left[2\,(\text{excess air factor})\right]$$

$$= 32{,}340 \text{ lb dry air/h (14.68 t/h)}$$

c. **Step 3.** Heat required to raise ambient air temperature
The basic formula for determining the heat required is:
Q = mass × heat capacity × temperature change
Heat required for raising the dry air temperature from 60°F (15.6°C) to 1400°F (760°C)

$$= \left(\frac{32{,}340 \text{ lb dry air}}{\text{h}}\right)\left(\frac{0.256 \text{ Btu}}{\text{lb} - {}^\circ\text{F}}\right)(1400^\circ\text{F} - 60^\circ\text{F})$$

$$= 11.09 \times 10^6 \text{ Btu/h (11.70 GJ/h)}$$

Heat required for raising the temperature of water vapor in air from 60°F (15.6°C) to 1400°F (760°C).

$$= \left(\frac{32{,}340 \text{ lb dry air}}{\text{h}}\right)\left(\frac{0.013 \text{ lb water}}{\text{lb air}}\right)\left(\frac{0.5 \text{ Btu}}{\text{lb} - {}^\circ\text{F}}\right)(1400^\circ\text{F} - 60^\circ\text{F})$$

$$= 0.28 \times 10^6 \text{ Btu/h (0.30 GJ/h)}$$

d. **Step 4.** Heat required to raise solids temperature
Heat required for raising the temperature of the volatile (combustible) material from 60°F (15.6°C) to 1400°F (760°C)

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.2 \text{ lb solids}}{\text{lb sludge}}\right)\left(\frac{0.77 \text{ lb VS}}{\text{lb solid}}\right)\left(\frac{0.25 \text{ Btu}}{\text{lb} - {}^\circ\text{F}}\right)$$

$$= 0.72 \times 10^6 \text{ Btu/h (0.76 GJ/h)}$$

Heat required for raising the temperature of inerts (ash) from 60°F (15.6°C) to the ash discharge temperature of 200°F (93.3°C)

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.2 \text{ lb solids}}{\text{lb sludge}}\right)\left(\frac{1 - 0.77 \text{ lb inerts}}{\text{lb solids}}\right)\left(\frac{0.25 \text{ Btu}}{\text{lb} - {}^\circ\text{F}}\right)(200^\circ\text{F} - 60^\circ\text{F})$$

$$= 0.02 \times 10^6 \text{ Btu/h (0.02 GJ/h)}$$

e. **Step 5.** Heat required to raise temperature of water associated with the feed sludge. This calculation does not include water formed during the combustion reaction. Heat required for raising the water temperature from 60°F (15.6°C) to 212°F (100°C)

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.8 \text{ lb water}}{\text{lb sludge}}\right)\left(\frac{1 \text{ Btu}}{\text{lb} - {}^\circ\text{F}}\right)(212^\circ\text{F} - 60^\circ\text{F})$$

$$= 1.70 \times 10^6 \text{ Btu/h (1.79 GJ/h)}$$

Heat required for evaporating water

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.8 \text{ lb water}}{\text{lb sludge}}\right)\left(\frac{970.3 \text{ Btu}}{\text{lb}}\right)$$

$$= 10.87 \times 10^6 \text{ Btu/h } (11.46 \text{ GJ/h})$$

Heat required for raising the temperature of water vapor to 1400°F (760°C)

$$= \left(\frac{14{,}000 \text{ lb sludge}}{\text{h}}\right)\left(\frac{0.8 \text{ lb water}}{\text{lb sludge}}\right)\left(\frac{0.5 \text{ Btu}}{\text{lb} - °\text{F}}\right)(1400°\text{F} - 212°\text{F})$$

$$= 6.65 \times 106 \text{ Btu/h } (7.02 \text{ GJ/h})$$

f. **Step 6.** Heat required to raise temperature of water formed during the combustion reaction. Assume water formed during the combustion reaction to be 0.5 lb/10,000 Btu (21.5 g/MJ) of sludge and supplemental fuel burned (21). The heat value of the sludge burned and supplemental fuel are equal to the heat demands. Therefore, water formed during combustion must be calculated on the basis of heat demands. Heat demands may be approximated by summing the calculations thus far:

Heat required	Btu/h × 10^6
Air	
Dry air	11.09
Water vapor in air	0.28
Sludge	
Volatile solids	0.72
Inerts	0.02
Free water	
Water	1.70
Evaporation	10.87
Water vapor	6.65
Total	31.33 (33.05 GJ/h)

Water formed resulting from the combustion reaction

$$= \left(\frac{0.5 \text{ lb}}{10{,}000 \text{ Btu}}\right)\left(\frac{31.33 \times 10^6 \text{ Btu}}{\text{h}}\right) = 1567 \text{ lb/h } (711 \text{ kg/h})$$

The heat of combustion given is the “high heat of combustion,” which assumes all water formed is condensed. Heat must be provided to evaporate this water and bring it up to exhaust temperature.

Heat required for evaporating the water

$$= \left(\frac{1567 \text{ lb water}}{\text{h}}\right)\left(\frac{970.3 \text{ Btu}}{\text{lb}}\right) = 1.52 \times 10^6 \text{ Btu/h } (1.60 \text{ GJ/h})$$

Heat required for raising the temperature of water vapor to 1400°F (760°C)

$$= \left(\frac{1567 \text{ lb water}}{\text{h}}\right)\left(\frac{0.5 \text{ Btu}}{\text{lb} - °\text{F}}\right)(1400°\text{F} - 212°\text{F}) = 0.93 \times 10^6 \text{ Btu/h } (0.98 \text{ GJ/h})$$

g. **Step 7.** Heat required to compensate for radiation losses

Assume a radiation loss of 5% of the total heat demand. Total heat demand is

Heat required	Btu/h × 10^6
Total from Step 6	31.33
Water formed during combustion	
Evaporation	1.52
Water vapor	0.93
Total	33.78 (35.64 GJ/h)

Heat to compensate for radiation losses

$$= \left(\frac{33.78 \times 10^6 \text{ Btu}}{\text{h}}\right)(0.05) = 1.69 \times 10^6 \text{ Btu/h (1.78 GJ/h)}$$

h. **Step 8.** Determine supplemental fuel required.
Total heat requirements (from Step 7)
= 33.78 × 10^6 Btu/h + 1.69 x 10^6 Btu/h
= 35.47 × 10^6 Btu/h (37.42 GJ/h)
Total supplemental heat demand
= Heat demand minus heating value of sludge
= (35.47 × 10^6 – 21.56 × 10^6) Btu/h
= 13.91 × 10^6 Btu/h (14.68 GJ/h)

Therefore, supplemental fuel (no. 2 fuel oil) must be supplied to provide 13.91 × 10^6 Btu/h (14.68 GJ/h) of heat. Supplemental fuel also requires air for combustion, and this air exerts a heat demand. The air required for supplemental fuel is 1.25 times the theoretical value needed for supplemental fuel.
Air required for supplemental fuel

$$= \left(\frac{13.91 \times 10^6 \text{ Btu}}{\text{h}}\right)\left(\frac{7.5 \text{ lb dry air}}{10{,}000 \text{ Btu}}\right)(1.25 \text{ [excess air factor]})$$

$$= 13{,}000 \text{ lb dry air/h (5920 kg/h)}$$

The 13,041 lb/h (5920 kg/h) dry air (plus any water formed by its reaction with the supplemental fuel) must also be raised to 1400°F (760°C). By calculations similar to those presented in Steps 1–8, it can be shown that the heat required to do this (and to account for additional radiation losses) is 20.24 × 10^6 Btu/h (21.35 GJ/h). Because only 13.91 × 10^6 Btu/h (14.68 GJ/h) was released by burning supplemental fuel, there is a heat deficit of 20.24 × 10^6 – 13.91 × 10^6 = 6.33 × 10^6 Btu/h (6.67 GJ/h). Thus, the effect of adding supplemental fuel was to reduce, but not eliminate, the initial deficit of 13.91 × 10^6 Btu/h (14.68 GJ/h). To make up for this deficit, more supplemental fuel, equivalent to 6.33 × 10^6 Btu/h (6.68 GJ/h) is added. If 25% excess air is used for this fuel, 5934 lb/h (2694 kg/h) of excess air will be required. The heat released is again insufficient to raise the air plus water vapor formed to 1400°F (760°C) and to make up for additional radiation losses. The deficit for this iteration is 2.88 × 10^6 Btu/h (3.04 GJ/h).

The calculation can be carried forward for several more steps. Table 10 shows that progressively smaller additions of supplemental fuel and air are required for each iteration and that the amount of air and fuel needed per iteration is a fixed fraction (0.45) of the fuel and air needed for the previous iteration. In general, if fuel required per iteration is r percent of that required for the previous iteration, then
Total fuel required = (initial deficit) $(1 + r + r^2 + r^3 + \cdots + r^n)$.

Table 10
Approximate Combustion Calculation—Supplemental Fuel Requirements

Heat input		Heat demands					
Unit	Heat value 10^6 Btu/h	Unit	Heat value 10^6 Btu/h	Supplemental fuel requirements (10^6 Btu/h)	Ratio[a]	Combustion air requirements (lb/h)	Ratio[b]
Sludge	21.56	Sludge and excess air	35.47	13.91	–	32,340	–
Supplemental fuel	13.91	Supplemental fuel and excess air	20.24	6.33	0.46	13,041	–[c]
Supplemental fuel	6.33	Supplemental fuel and excess air	9.21	2.88	0.45	5934	0.46
Supplemental fuel	2.88	Supplemental fuel and excess air	4.19	1.31	0.45	2700	0.46
Supplemental fuel	1.31	Supplemental fuel and excess air	1.91	0.60	0.46	1228	0.45

[a] Ratio of supplemental fuel to that in the previous iteration.
[b] Ratio of air to air in the previous iteration.
[c] Ratio in this case is not applicable since sludge is included (100% excess air vs 25% excess air).
1 × 10^6 Btu/h = 1055 MJ/h.
1 lb/h = 0.45 kg/h.

The term in the second bracket is an infinite geometric series equal to r^n. The series converges to $1/(1–r)$, if the absolute value of $r < 1$ (22).

The total supplemental fuel requirements can be derived from Eq. (4).

$$\text{Total supplemental fuel} = (\text{Initial deficit})\left(\frac{1}{1-r}\right) \tag{4}$$

$$\text{Total supplemental fuel} = \left(13.91 \times 10^6 \ \frac{\text{Btu}}{\text{h}}\right)\left(\frac{1}{1-0.45}\right)$$
$$= 25.32 \times 10^6 \ \text{Btu/h (26.6 GJ/h)}$$

i. **Step 9.** Total air requirements
The air requirements for the supplemental fuel alone can be found from Eq. (5), an analog to Eq. (4).
Total supplemental air requirements

$$= \text{excess air for initial supplemental fuel addition}\left(\frac{1}{1-r}\right) \tag{5}$$

$$= \left(\frac{13{,}041 \text{ lb air}}{\text{h}}\right)\left(\frac{1}{1-0.45}\right) = 23{,}735 \text{ lb dry air/h } (10{,}766 \text{ kg dry air/h})$$

Total dry air requirements
= air for sludge plus air for supplemental fuel (32,340 + 23,735) lb dry air/h
= 56,075 lb dry air/h (25,458 kg/h)
Assuming an air density of 0.0749 lb/ft^3 (1.2 kg/m^3):
Air flow rate

$$= \left(\frac{56{,}075 \text{ lb/h}}{0.0749 \text{ lb/ft}^3}\right)\left(\frac{1 \text{ h}}{60 \text{ min}}\right) = 12{,}478 \text{ ft}^3\text{/min } (5.9 \text{ m}^3\text{/s})$$

Assume that no. 2 oil has heating value of 141,000 Btu/gal
Supplemental fuel rate

$$= \left(\frac{25.32 \times 10^6 \text{ Btu/h}}{141{,}000 \text{ Btu/gal}}\right)\left(\frac{1 \text{ h}}{60 \text{ min}}\right) = 3 \text{ gal/min } (0.18 \text{ l/s})$$

4.3. Theoretical Calculation Method

The method presented here is based on the actual combustion reactions and the method of approach used in steam generation calculations (21). Table 11 is to be used for calculations as follows (lines numbered on left and alphabetical on right):

a. **Step 1 (Line b).** Determine the fuel analysis and include on the right-hand side of the table (ultimate analysis).
b. **Step 2 (Lines 1–12).** Determine the pounds of component, moles of component, theoretical oxygen requirement and moles of material contributed to the fuel gas by the fuel, based on 100 lb of fuel feed. Assume complete combustion and no loss of combustibles to the ash.
c. **Step 3. (Lines 13–15).** Assume the amount of excess O_2 to be used (100%) and calculate the moles of excess O_2 required.
d. **Step 4 (Line 16).** Calculate the amount of N_2 added from the air from the total O_2 (theoretical plus excess).
e. **Step 5 (Lines 17–19 and 21).** Calculate the amount of dry air, water in the air, the amount of wet air from the total dry air (O_2 + N_2).
f. **Step 6 (Lines 20 and s).** Calculate the moles of all components in the fuel gas and the moles of wet and dry fuel gas.
g. **Step 7 (Lines 22–26).** Determine the sensible heat content of the gas. A base temperature of 60°F (15°C) is used. The values for mean specific heat can be found in ref. 23. Note that mean molar specific heat = mean specific heat × molecular weight.
h. **Step 8 (Line 27).** Determine the latent heat of water in the fuel gas.
i. **Step 9 (Line 28).** Sum all heat in fuel gas.
j. **Step 10 (Lines 29–31).** Calculate heat losses owing to carbon in refuse (residue), unburned CO in the fuel gas, and radiation (assumed to be 5%). Sum all heat losses.
k. **Step 11 (Line 32).** Determine heat value of the sludge per 100 lb, wet basis.
l. **Step 12 (Line 33).** Determine if the sludge is autogenous or requires supplemental fuel by subtracting line 32 from line 31. A zero or positive number indicates that the sludge is autogenous, supplemental fuel is not required, and the computation is complete. A negative number shows that supplemental fuel is necessary. The method used to determine the amount of fuel required is shown in steps 13–15.
m. **Step 13.** If Step 12 indicates that supplemental fuel is required, proceed through another theoretical calculation method table for the supplemental fuel in the same manner as Steps 1–12 (lines 1–33). This determines the amount of excess heat in the fuel after the combustion reaction. Table 12 illustrates the supplemental fuel calculation for this example.

Table 11
Combustion Calculations—Molal Basis

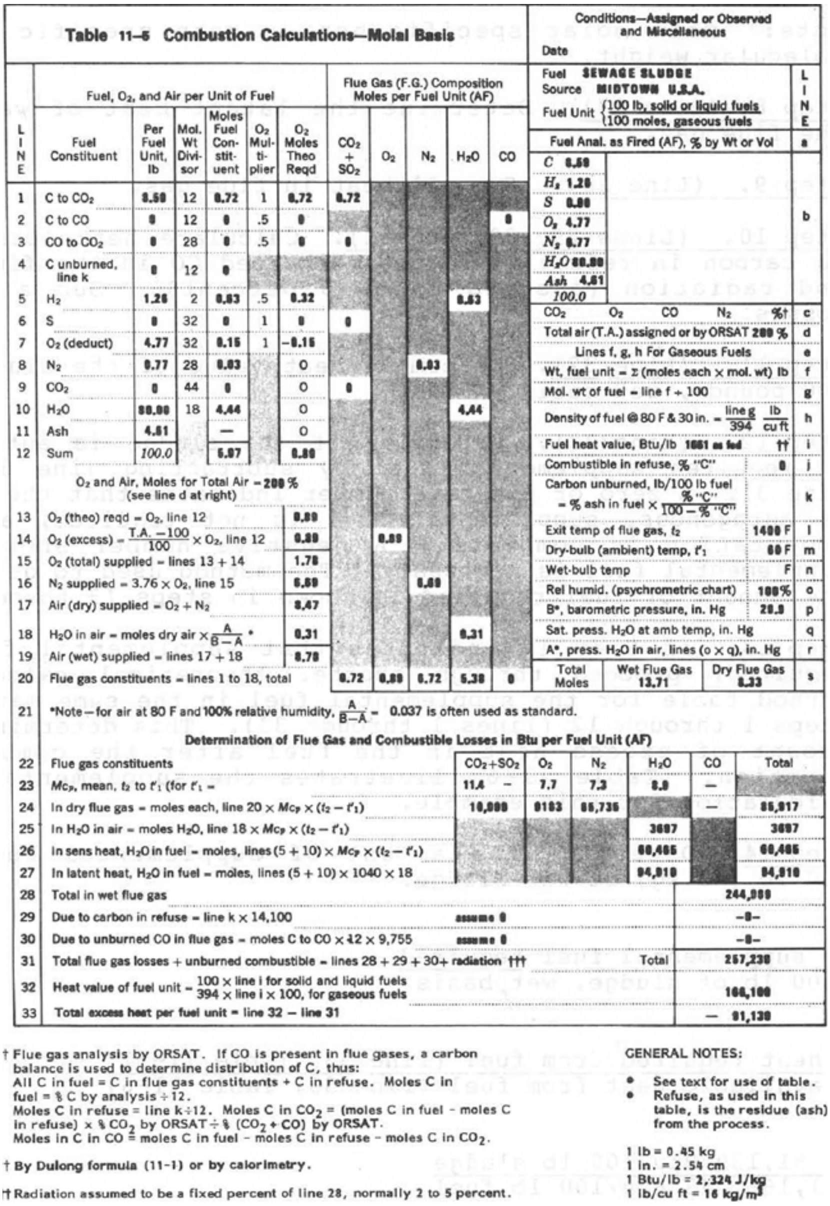

Table 11–5 Combustion Calculations—Molal Basis

Fuel, O_2, and Air per Unit of Fuel / Flue Gas (F.G.) Composition Moles per Fuel Unit (AF)

LINE	Fuel Constituent	Per Fuel Unit, lb	Mol. Wt Divisor	Moles Fuel Constituent	O_2 Multiplier	O_2 Moles Theo Reqd	CO_2 + SO_2	O_2	N_2	H_2O	CO
1	C to CO_2	8.60	12	0.72	1	0.72	0.72				
2	C to CO	0	12	0	.5	0					0
3	CO to CO_2	0	28	0	.5	0	0				
4	C unburned, line k	0	12	0							
5	H_2	1.26	2	0.63	.5	0.32				0.63	
6	S	0	32	0	1	0	0				
7	O_2 (deduct)	4.77	32	0.15	1	−0.15					
8	N_2	0.77	28	0.03		0			0.03		
9	CO_2	0	44	0		0	0				
10	H_2O	80.00	18	4.44		0				4.44	
11	Ash	4.61		—		0					
12	Sum	100.0		5.97		0.89					
	O_2 and Air, Moles for Total Air = 200 % (see line d at right)										
13	O_2 (theo) reqd = O_2, line 12					0.89					
14	O_2 (excess) = $\frac{T.A. - 100}{100} \times O_2$, line 12					0.89		0.89			
15	O_2 (total) supplied = lines 13 + 14					1.78					
16	N_2 supplied = 3.76 × O_2, line 15					6.69			6.69		
17	Air (dry) supplied = $O_2 + N_2$					8.47					
18	H_2O in air = moles dry air × $\frac{A}{B-A}$ *					0.31				0.31	
19	Air (wet) supplied = lines 17 + 18					8.78					
20	Flue gas constituents = lines 1 to 18, total						0.72	0.89	6.72	5.38	0
21	*Note—for air at 80 F and 100% relative humidity, $\frac{A}{B-A}$ = 0.037 is often used as standard.										

Conditions—Assigned or Observed and Miscellaneous

Date

Fuel **SEWAGE SLUDGE**
Source **MIDTOWN U.S.A.**
Fuel Unit {100 lb, solid or liquid fuels / 100 moles, gaseous fuels

		LINE
Fuel Anal. as Fired (AF), % by Wt or Vol		a
C 8.60 H_2 1.26 S 0.00 O_2 4.77 N_2 0.77 H_2O 80.00 Ash 4.61 100.0		b
CO_2 O_2 CO N_2	%†	c
Total air (T.A.) assigned or by ORSAT	200 %	d
Lines f, g, h For Gaseous Fuels		e
Wt, fuel unit = Σ (moles each × mol. wt) lb		f
Mol. wt of fuel = line f ÷ 100		g
Density of fuel @ 80 F & 30 in. = $\frac{\text{line g}}{394}$ $\frac{\text{lb}}{\text{cu ft}}$		h
Fuel heat value, Btu/lb 1661 as fed	††	i
Combustible in refuse, % "C"	0	j
Carbon unburned, lb/100 lb fuel = % ash in fuel × $\frac{\% \text{"C"}}{100 - \% \text{"C"}}$		k
Exit temp of flue gas, t_2	1400 F	l
Dry-bulb (ambient) temp, t'_1	80 F	m
Wet-bulb temp	F	n
Rel humid. (psychrometric chart)	100%	o
B*, barometric pressure, in. Hg	29.9	p
Sat. press. H_2O at amb temp, in. Hg		q
A*, press. H_2O in air, lines (o × q), in. Hg		r
Total Moles: Wet Flue Gas 13.71	Dry Flue Gas 8.33	s

Determination of Flue Gas and Combustible Losses in Btu per Fuel Unit (AF)

LINE		CO_2+SO_2	O_2	N_2	H_2O	CO	Total
22	Flue gas constituents						
23	Mc_p, mean, t_2 to t'_1 (for t'_1 =	11.4 –	7.7	7.3	9.0	—	
24	In dry flue gas = moles each, line 20 × Mc_p × $(t_2 - t'_1)$	10,998	9183	65,736		—	85,917
25	In H_2O in air = moles H_2O, line 18 × Mc_p × $(t_2 - t'_1)$				3687		3687
26	In sens heat, H_2O in fuel = moles, lines (5 + 10) × Mc_p × $(t_2 - t'_1)$				60,465		60,465
27	In latent heat, H_2O in fuel = moles, lines (5 + 10) × 1040 × 18				94,910		94,910
28	Total in wet flue gas						244,989
29	Due to carbon in refuse = line k × 14,100	assume 0					-0-
30	Due to unburned CO in flue gas = moles C to CO × 12 × 9,755	assume 0					-0-
31	Total flue gas losses + unburned combustible = lines 28 + 29 + 30+ radiation †††					Total	257,239
32	Heat value of fuel unit = 100 × line i for solid and liquid fuels / 394 × line i × 100, for gaseous fuels						166,100
33	Total excess heat per fuel unit = line 32 − line 31						— 91,139

† Flue gas analysis by ORSAT. If CO is present in flue gases, a carbon balance is used to determine distribution of C, thus:
All C in fuel = C in flue gas constituents + C in refuse. Moles C in fuel = % C by analysis ÷ 12.
Moles C in refuse = line k÷12. Moles C in CO_2 = (moles C in fuel − moles C in refuse) × % CO_2 by ORSAT ÷ % (CO_2 + CO) by ORSAT.
Moles in C in CO = moles C in fuel − moles C in refuse − moles C in CO_2.

† By Dulong formula (11-1) or by calorimetry.

†† Radiation assumed to be a fixed percent of line 28, normally 2 to 5 percent.

GENERAL NOTES:

- See text for use of table.
- Refuse, as used in this table, is the residue (ash) from the process.

1 lb = 0.45 kg
1 in. = 2.54 cm
1 Btu/lb = 2,324 J/kg
1 lb/cu ft = 16 kg/m³

n. **Step 14.** Determine the amount of supplemental fuel per 100 lb (45 kg) of wet sludge.

$$\frac{\text{lb supplemental fuel required}}{\text{100 lb of sludge (wet basis)}} = \frac{\text{heat required from fuel (line 33, Table 11-5)}}{\text{available heat from fuel (line 33, Table 11-6)}}$$

$$\frac{\text{91,139 Btu/100 lb sludge}}{\text{1,165,443 Btu/100 lb fuel}} = \text{7.82 lb fuel/100 lb sludge (wet basis).}$$

Table 12
Combustion Calculations—Molal Basis

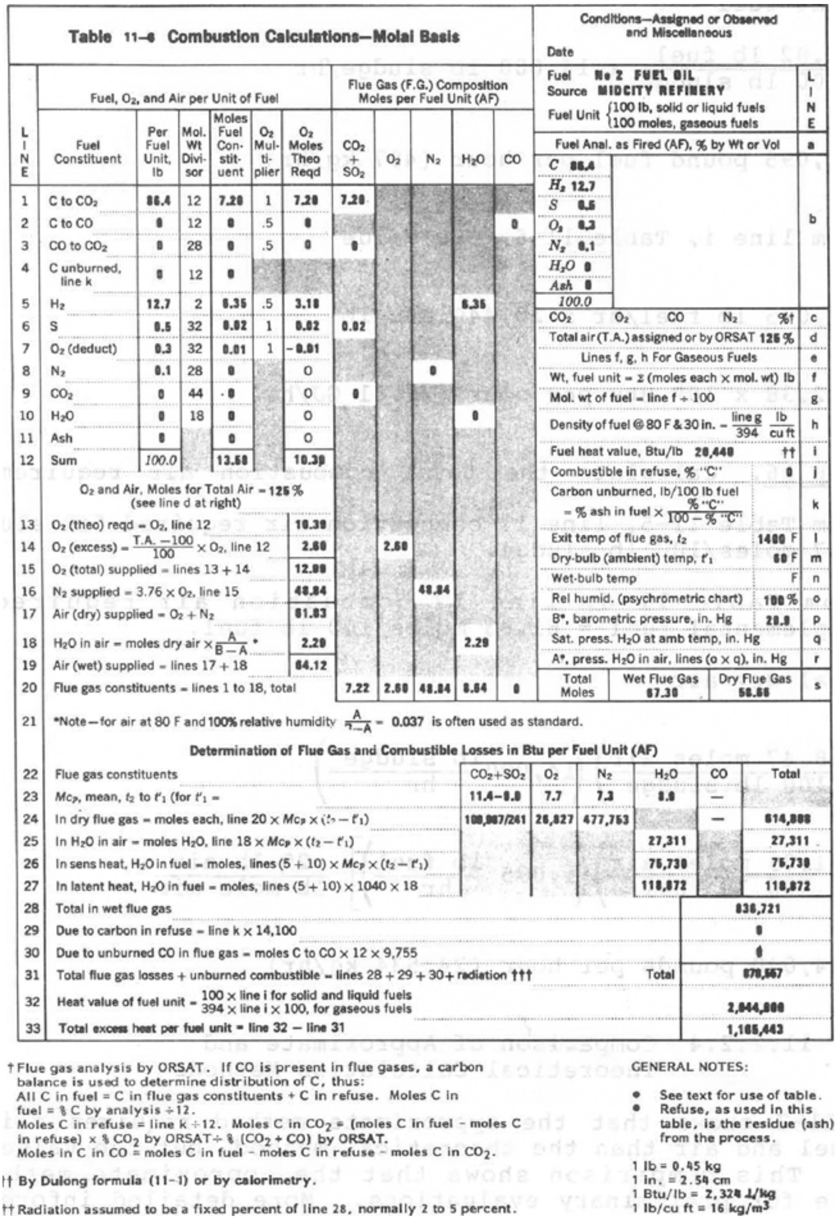

Table 11–6 Combustion Calculations—Molal Basis

LINE	Fuel Constituent	Per Fuel Unit, lb	Mol. Wt Divisor	Moles Fuel Constituent	O_2 Multiplier	O_2 Moles Theo Reqd	CO_2 + SO_2	O_2	N_2	H_2O	CO
	Fuel, O_2, and Air per Unit of Fuel						**Flue Gas (F.G.) Composition Moles per Fuel Unit (AF)**				
1	C to CO_2	86.4	12	7.20	1	7.20	7.20				
2	C to CO	0	12	0	.5	0					0
3	CO to CO_2	0	28	0	.5	0	0				
4	C unburned, line k	0	12	0							
5	H_2	12.7	2	6.35	.5	3.18				6.35	
6	S	0.5	32	0.02	1	0.02	0.02				
7	O_2 (deduct)	0.3	32	0.01	1	−0.01					
8	N_2	0.1	28	0		0			0		
9	CO_2	0	44	·0		0	0				
10	H_2O	0	18	0		0				0	
11	Ash	0		0		0					
12	Sum	100.0		13.58		10.39					
	O_2 and Air, Moles for Total Air = 125% (see line d at right)										
13	O_2 (theo) reqd = O_2, line 12					10.39					
14	O_2 (excess) = $\frac{T.A. - 100}{100} \times O_2$, line 12					2.60		2.60			
15	O_2 (total) supplied = lines 13 + 14					12.99					
16	N_2 supplied = 3.76 × O_2, line 15					48.84			48.84		
17	Air (dry) supplied = O_2 + N_2					61.83					
18	H_2O in air = moles dry air × $\frac{A}{B-A}$ *					2.29				2.29	
19	Air (wet) supplied = lines 17 + 18					64.12					
20	Flue gas constituents = lines 1 to 18, total						7.22	2.60	48.84	8.64	0

		LINE
Conditions—Assigned or Observed and Miscellaneous		
Date		
Fuel **No 2 FUEL OIL** Source **MIDCITY REFINERY** Fuel Unit {100 lb, solid or liquid fuels; 100 moles, gaseous fuels}		
Fuel Anal. as Fired (AF), % by Wt or Vol		a
C 86.4 *H_2* 12.7 *S* 0.5 *O_2* 0.3 *N_2* 0.1 *H_2O* 0 *Ash* 0 *100.0*		b
CO_2 O_2 CO N_2	%†	c
Total air (T.A.) assigned or by ORSAT	125 %	d
Lines f, g, h For Gaseous Fuels		e
Wt, fuel unit = Σ (moles each × mol. wt) lb		f
Mol. wt of fuel = line f ÷ 100		g
Density of fuel @ 80 F & 30 in. = $\frac{\text{line g}}{394}$ $\frac{\text{lb}}{\text{cu ft}}$		h
Fuel heat value, Btu/lb 20,440	††	i
Combustible in refuse, % "C"	0	j
Carbon unburned, lb/100 lb fuel = % ash in fuel × $\frac{\%\ \text{"C"}}{100 - \%\ \text{"C"}}$		k
Exit temp of flue gas, t_2	1400 F	l
Dry-bulb (ambient) temp, t'_1	60 F	m
Wet-bulb temp	F	n
Rel humid. (psychrometric chart)	100 %	o
B*, barometric pressure, in. Hg	29.9	p
Sat. press. H_2O at amb temp, in. Hg		q
A*, press. H_2O in air, lines (o × q), in. Hg		r
Total Moles	Wet Flue Gas 67.30 / Dry Flue Gas 58.66	s

21 *Note—for air at 80 F and 100% relative humidity $\frac{A}{B-A}$ = 0.037 is often used as standard.

Determination of Flue Gas and Combustible Losses in Btu per Fuel Unit (AF)

LINE		CO_2+SO_2	O_2	N_2	H_2O	CO	Total
22	Flue gas constituents	CO_2+SO_2	O_2	N_2	H_2O	CO	Total
23	Mc_p, mean, t_2 to t'_1 (for t'_1 =	11.4–9.0	7.7	7.3	8.9	—	
24	In dry flue gas = moles each, line 20 × Mc_p × ($t_2 - t'_1$)	109,987/241	26,827	477,753		—	614,808
25	In H_2O in air = moles H_2O, line 18 × Mc_p × ($t_2 - t'_1$)				27,311		27,311
26	In sens heat, H_2O in fuel = moles, lines (5 + 10) × Mc_p × ($t_2 - t'_1$)				75,730		75,730
27	In latent heat, H_2O in fuel = moles, lines (5 + 10) × 1040 × 18				118,872		118,872
28	Total in wet flue gas						836,721
29	Due to carbon in refuse = line k × 14,100						0
30	Due to unburned CO in flue gas = moles C to CO × 12 × 9,755						0
31	Total flue gas losses + unburned combustible = lines 28 + 29 + 30 + radiation †††					Total	878,557
32	Heat value of fuel unit = 100 × line i for solid and liquid fuels 394 × line i × 100, for gaseous fuels						2,044,000
33	Total excess heat per fuel unit = line 32 − line 31						1,165,443

† Flue gas analysis by ORSAT. If CO is present in flue gases, a carbon balance is used to determine distribution of C, thus:
All C in fuel = C in flue gas constituents + C in refuse. Moles C in fuel = % C by analysis ÷ 12.
Moles C in refuse = line k ÷ 12. Moles C in CO_2 = (moles C in fuel − moles C in refuse) × % CO_2 by ORSAT ÷ % (CO_2 + CO) by ORSAT.
Moles in C in CO = moles C in fuel − moles C in refuse − moles C in CO_2.

†† By Dulong formula (11–1) or by calorimetry.

†† Radiation assumed to be a fixed percent of line 28, normally 2 to 5 percent.

GENERAL NOTES:

- See text for use of table.
- Refuse, as used in this table, is the residue (ash) from the process.

1 lb = 0.45 kg
1 in. = 2.54 cm
1 Btu/lb = 2,324 J/kg
1 lb/cu ft = 16 kg/m^3

o. **Step 15.** Calculate the total fuel demand for 14,000 lb/h of wet sludge (6356 kg/h):

$$\text{Total fuel} = \left(\frac{7.82\ \text{lb fuel}}{100\ \text{lb sludge}}\right)\left(\frac{14{,}000\ \text{lb sludge}}{\text{h}}\right) = 1095\ \text{lb fuel/h (497 kg/h)}$$

From line i, Table 12, Btu value

$$= \left(\frac{1095\ \text{lb fuel}}{\text{h}}\right)\left(\frac{20{,}440\ \text{Btu}}{\text{lb}}\right) = 22.38\ \times\ 106\ \text{Btu/h (23.61 GJ/h)}$$

Table 13
Comparison Between An Approximate and a Theoretical Calculation of Furnace Combustion

	AM		TM		
Item	Value	Calculation reference (AM)	Value	Calculation reference (TM)	Difference (AM–TM)/TM × 100
Sludge heating value	$\frac{10{,}000\ \text{Btu}^{a}}{\text{lb VS}}$	Assumed	$\frac{1661\ \text{Btu}^{a}}{\text{lb as fed}}$	Table 11–4 line i	–7.28
Furnace heat deficit	$13.91 \times 10^{6}\ \frac{\text{Btu}}{\text{h}}$	Step 8	$\frac{91{,}139\ \text{Btu}^{b}}{100\ \text{lb wet sludge}}$	Table 11–4 line 33	9.01
Supplemental fuel heating value	$\frac{141{,}000\ \text{Btu}^{c}}{\text{gal}}$	Step 8	$\frac{20{,}440\ \text{Btu}}{\text{lb}}$	Table 11–5 line i	–4.19
Supplemental fuel required	$25.32 \times 10^{6}\ \frac{\text{Btu}}{\text{h}}$	Step 8	$22.38 \times 10^{6}\ \frac{\text{Btu}}{\text{h}}$	Step 15	13.14
Total combustion air required	$56{,}075\ \frac{\text{lb}}{\text{h}}$	Step 9	$54{,}040\ \frac{\text{lb}}{\text{h}}$	Step 16	3.77

AM, approximate method; TM, theoretical method.

[a] 10,000 Btu/lb VS at 77% VS = 7700 Btu/lb dry solids, 1661 Btu/lb as fed t 20% solids = 8305 Btu/lb dry solids.

[b] 91,139 Btu/100 Ib wet × 14,000 Ib wet/h = 12.76 × 106 Btu/h.

[c] 141,000 Btu/gal; 7.2 Ib/gal = 19,583 Btu/lb.

1 Btu/lb = 2324 J/kg.
1 Btu/h = 0.293 kw.
1 Btu/gal = 279 J/L.
1 Ib/h = 0.45 kg/h.

p. **Step 16.** Calculate the total combustion air requirements:
From Table 11, line 17 combustion air required for sludge = 8.47 moles/100 lb sludge.
From Table 12, line 17 combustion air required for supplemental fuel = 61.83 moles/l00 lb fuel.
Total dry air

$$= \left(\left[\frac{8.47\ \text{moles air}}{1001\text{b sludge}} \right] \left[\frac{14{,}000\ \text{lb sludge}}{\text{h}} \right] + \left[\frac{61.83\ \text{moles air}}{100\ \text{lb fuel}} \right] \left[\frac{1{,}095\ \text{lb fuel}}{\text{h}} \right] \right) \left(\frac{29\ \text{lb air}}{\text{lb mole air}} \right)$$

= 54,040 lb/h (24,534 kg/h)

4.4. *Comparison of Approximate and Theoretical Calculation Methods*

Table 13 shows that the approximate method (AM) requires slightly more fuel and air than the TM, but the values are close. This comparison shows that the AM is suitable for preliminary evaluations. More detailed information and combustion theory can be found in the literature (1,23,6,7,21, and 24–29).

5. REGULATORY MATTERS

The incineration of sludge is regulated under 40 CFR Part 503. Part 503 is the comprehensive regulation for land application, surface disposal, and incineration of sewage sludge. Round-1 regulations were published in February 1993, and have been amended several times since then. In December 2001, a notice was given by the Administrator (US EPA) that numerical standards and management practices are not warranted for dioxins in sewage sludge disposed off in a surface disposal unit or fired in a sewage sludge incinerator (30).

NOMENCLATURE

C	The weight fraction of carbon determined by ultimate analysis (fraction)
H_2	The weight fraction of hydrogen determined by ultimate analysis (fraction)
O_2	The weight fraction of oxygen determined by ultimate analysis (fraction)
S	The weight fraction of sulfur determined by ultimate analysis (fraction)
P	Minimum percent dry solids in sludge required for autogenous combustion (%)
Q	Fuel value of sludge (Btu/lb of dry solids)
W	Heat required to evaporate 1 lb of water in a multiple hearth furnace (Btu)
r	Fuel required per iteration (% of that required for the previous iteration)

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Biosolids Composting

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CONTENTS

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 increased attention has been directed to composting of wastewater biosolids as an environmentally sound alternative to stabilization for biosolids reuse or disposal.

From: *Handbook of Environmental Engineering, Volume 6: Biosolids Treatment Processes*
Edited by: L. K. Wang et al. © The Humana Press Inc., Totowa, NJ

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, MD (3–5), and full-scale operations at County Sanitation Districts of Los Angeles County plant at Carson, CA. 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 enhanced the acceptance of biosolids as a resource by standardizing metal and pathogen concentrations. Moreover, Part 503 officially identifies composting as a method to control pathogens and reduce vector attraction (9).

In 1983 the number of composting facilities in the United States numbered only 61 but by 1988 the number of biosolids composting plants had risen to 115 (10,11). According to 1998 survey in *Biocycle, The Journal of Composting and Recycling*, 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 (>40%) 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 inconveniences such as odor, 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 principle advantages of biosolids composting as compared with refuse composting (15), which are as follows:

a. Biosolids composting does not require the complex materials management and separation techniques necessary for most refuse composting operations.
b. Municipal wastewater biosolids are more uniform in composition causing less operating difficulties.
c. The final composted mixture utilizing biosolids is more suitable for marketing because it generally does not contain plastics, metal, and glass commonly found in refuse compost.
d. 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 which are as follows:

a. Windrow.
b. Aerated static pile.
c. In-vessel systems.

Each method uses the same scientific principles 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), which are as follows:

a. Lack of availability of landfill space for solids disposal.
b. Composting economics are more favorable when landfill tipping fees escalate.
c. Emphasis on beneficial reuse at federal, state, and local levels.
d. Ease of storage, handling, and use of composted product.
e. 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, it decomposes slowly, and remains effective for a longer time than the organic matter in uncomposted waste. Composted biosolids can improve the quality of soils containing excessive amount 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):

a. The recycling of a valuable resource.
b. Reduction of dependence on chemical fertilizers.
c. Offsetting the use of natural resources such as trees or peat moss as mulch material.
d. Provides organic nitrogen, phosphorus, and potassium.
e. Provides essential plant micronutrients.
f. Can reduce the need for pesticides.
g. Increased water holding of soils.
h. Increased aeration and drainage for clay soils.
i. Increased permeability for clay soils.
j. Greater root depth.
k. Increased microbial population.
l. 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 water. 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) which are as follows:

a. Odor production at the composting site.
b. Survival and presence of primary pathogens in the product.
c. Dispersion of secondary pathogens such as *Aspergillus fumigatus*, particulate matter, other airborne allergens.
d. Lack of consistency in product quality with reference to metals, stability, and maturity.

Odor from a composting operation can be a nuisance and a potential irritant. Offensive odor 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 odor may not pose a health threat, odor 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 odor and protect nearby residents from exposure to malodor.

Composting odor 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 Fig. 4, operated at a high positive redox potential (highly aerobic) will reduce, but not necessarily eliminate, odor and odor causing compounds during the first 10–14 d of the process (23,24).

In addition to odor, other bioaerosols, such as pathogens, endotoxins, and various volatile organic compounds, must also be controlled. Biofilters are often used to control odor, 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 nondetectable levels if the temperature of the compost is maintained at more than 55 C for 15 d or more (25–27). Additionally, it has been demonstrated that viruses and helminth ova do not regrow after thermal inactivation (25).

Regrowth of *Salmonella* spp. 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 postcomposting handling and addition of amendments coupled with 4–6 wk of storage will mitigate *Salmonella* regrowth (25).

Compost workers may be exposed to a common fungus known as *A. fumigatus*, 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):

a. Wearing masks and other protective devices.
b. Equipping front end loaders with filters or air conditioners.
c. Thoroughly ventilating composting halls.
d. 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 who 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 US EPA (9), defines two types of biosolids with respect to pathogen reduction: classes A and B. The difference is defined by the degree of pathogen reduction on the solids. When federal performance standards are met, composting ensures full destruction of pathogens to *nondetectable levels* in the wastewater solids (i.e., to class A standards).

The period 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 classes A and B products (Table 1). The production of a class B product is not always economically justified because 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 concentration for nine metals which cannot be exceeded in biosolids products, including compost. These are known as ceiling concentrations. The federal maximum allowable metals concentration is provided in Table 2. The Part 503

Table 1
Time and Temperature Requirements for Biosolids Composting

Product	Regulatory requirements
Class A	Aerated static pile or in-vessel: 55°C for at least 3 d Windrow: 55°C for at least 15 d with 5 turns
Class B	About 40°C or higher for 5 d during which temperature exceed 55°C for at least 4 h

Source: From refs. 1,9.

Table 2
Maximum Metal Concentrations in Biosolids

Metal	Ceiling concentration (mg/kg)	Pollutant concentration (mg/kg)
Arsenic	75	41
Cadmium	85	39
Copper	4300	1500
Lead	840	300
Mercury	57	17
Molybdenum	75	NL
Nickel	420	420
Selenium	100	100
Zinc	7500	2800

Source: From refs. 9,18.
NL, no established limit.

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 more 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 preprocessed wastewater residuals, enabling compost to meet the stringent EQ standards of Part 503.

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 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 50–80 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 concentration, 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 foul odor (23,32).

4. PROCESS DESCRIPTION

The basic composting process consists of the following steps (15), which are as follows:

a. 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 (e.g., recycled compost, wood chips, and so on.) or feed amendments for a source of limiting nutrients such as carbon (for example, sawdust, rice hulls, and so on.) are added to the dewatered biosolids to provide a mixture suitable for composting.
b. Temperature in the range of 55–65°C (130–150°F) is required to ensure destruction of pathogenic organisms and provides the driving force for evaporation, which reduces the moisture content.
c. The compost is stored for extended period after the primary composting operation to further stabilize the mixture at lower temperatures.
d. Additional air-drying may be required if the cured compost is too wet for further processing.
e. 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 d 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 ensure 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 population of bacteria, actinomycetes, and other fungi associated with a diverse succession of environments. The principal factors which affect the microbiology of composting include (33), which are as follows:

a. Moisture.
b. Temperature.
c. pH.
d. Nutrient concentration.
e. 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%; however, less than 40% moisture may limit the rate of decomposition. The optimum moisture content is in

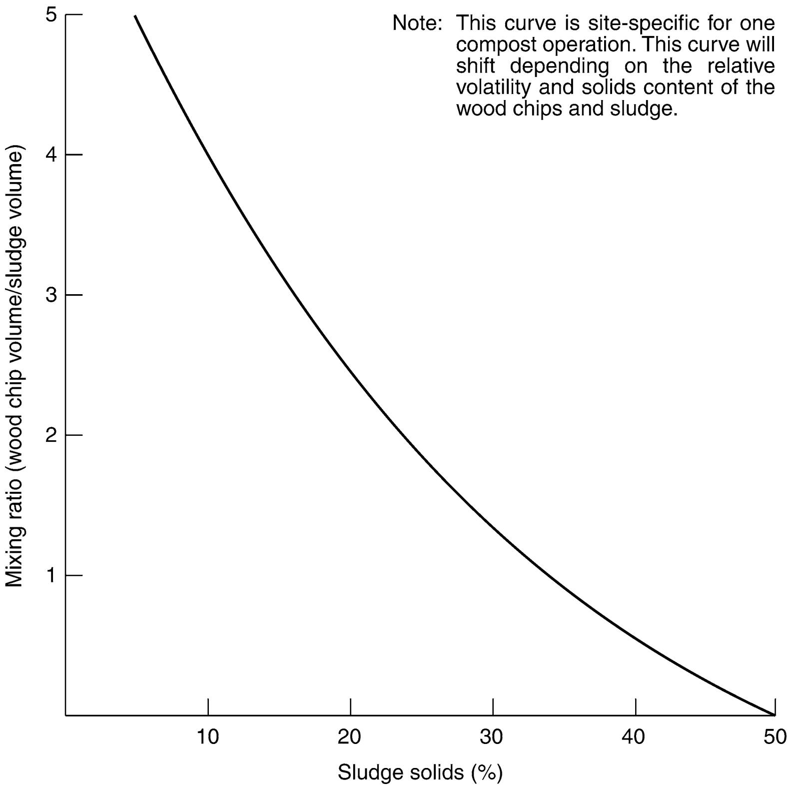


Fig. 1. Effect of solids content on the ratio of wood chips to biosolids by volume (15).

the range of 50–60%. If the mixture is over 60% 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 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% filter cake would be about one-fifth the amount required for a 20% solids cake. In addition to savings on wood chips, there would be a substantial reduction in material management costs and site sizes (34).

The US Composting Council (35) lists the following materials as suitable for use as bulking agents:

a. Agricultural byproducts, such as manure and bedding from various animals, animal mortalities, and crop residues.
b. Yard trimmings, including grass clippings, leaves, weeds, stumps, twigs, tree prunings, Christmas trees, and other vegetative matter from land clearing activities.

c. Food byproducts, including damaged fruits and vegetables, coffee grounds, peanut hulls, egg shells, and fish residues.
d. Industrial byproducts from wood processing, forestry, brewery and pharmaceutical operations. Paper goods, paper mill residues, and biodegradable packaging materials are also used.
e. Municipal solid waste.

4.2. Temperature

For the most efficient operation, the temperature in the compost should range between 55 and 65°C (130–150°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), the following factors:

a. Moisture content.
b. Aeration rates.
c. Size and shape of pile.
d. Atmospheric conditions.
e. 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, the 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 limits (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% by volume (37). Increasing the oxygen concentration beyond 15% by air addition will result in a temperature decrease because of the greater air flow. Although oxygen concentrations as low as 0.5% have been observed inside windrows without anaerobic symptoms, generally at least 5% oxygen is generally required for aerobic conditions (33).

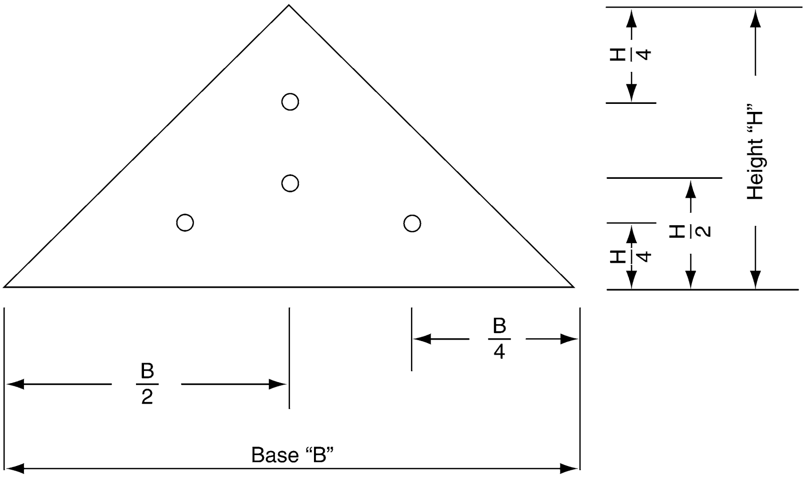


Fig. 2. Locations for temperature and oxygen monitoring at one end of a windrow or individual aerated pile (15).

5. DESIGN CRITERIA AND PROCEDURES

The basic criteria for successful composting is that the material to be composted be porous, and structurally stable and contain sufficient degradable material so that the degradation reaction is self-sustaining (i.e., 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. 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, 0–25% 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 (0–100°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 3. Four locations for temperature and oxygen measurements at both ends of each pile are shown in Fig. 2.

Haug and Haug (39) have shown the compost reaction is self-sustaining when the ratio W is ≤ 10: where W is the mass of water in compost mixture/mass of organics degraded by composting. In windrow and mechanical composting, porosity and structural stability

Table 3
Monitoring Program for a Municipal Wastewater Biosolids Composting Facility

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 5–15% oxygen content for 3–5 d	Temperature and oxygen content measurements taken at least 6 d during first 2 wk (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
	Odor	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
	Learhate 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

Source: From ref. 15.

are provided when the biosolids are mixed with recycled compost product or bulking agent to obtain solids concentration of approx 40–60%. 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, generally the bulking agents are screened out of the compost and recycled back to the mixing 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 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 Eqs. (1) and (2) derived from Fig. 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 Eq. (2) is used to determine the ratio *W* (39):

$$X_R = \frac{X_C\,(S_M - S_C) + X_A\,(S_M - S_A) + X_B\,(S_M - S_B)}{(S_R - S_M)} \tag{1}$$

$$W = \frac{X_C\,(1 - S_C) + X_A\,(1 - S_A) + X_B\,(1 - S_B) + X_R\,(1 - S_R)}{X_C\,S_C\,V_C\,k_C + X_A\,S_A\,V_A\,k_A + X_B\,S_B\,V_B\,k_B + X_R\,S_R\,V_R\,k_R} \tag{2}$$

5.1. Compost Processes With No External Bulking Agent

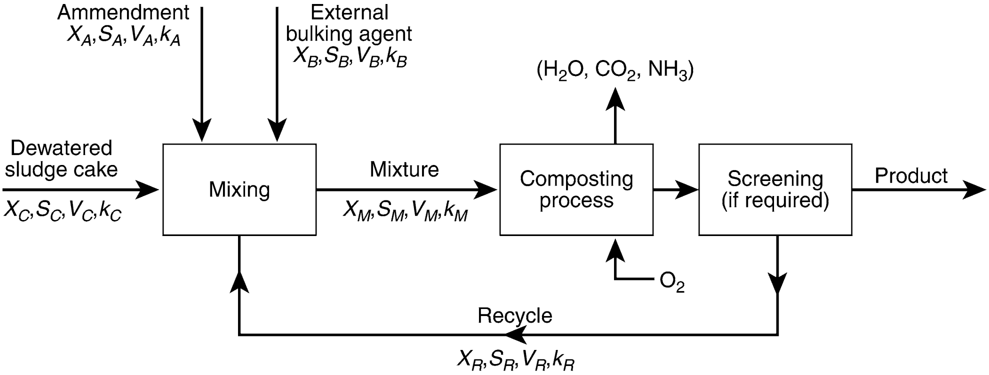
To design a compost facility employing no external bulking agent, the parameters X_C, S_C, V_C, k_C, S_R, V_R, k_R, and S_M must be determined analytically, assumed, or calculated. The wet weight of recycled compost (X_R) is calculated, assuming no amendment or external bulking agent addition ($X_A = X_B = O$), to provide a desired solids content of the mixture (S_M) in the 0.40–0.50 range:

$$X_R = \frac{X_C\,(S_M - S_C)}{(S_R - S_M)} \tag{3}$$

Once X_R is determined for these conditions, the ratio *W* is calculated:

$$W = \frac{X_C\,(1 - S_C) + X_R\,(1 - S_R)}{X_C\,S_C\,V_C\,k_C + X_R\,S_R\,V_R\,k_R} \tag{4}$$

If the ratio *W* is <10, the compost mixture has sufficient energy available for temperature elevation and water evaporation. The ratio number of 10 is not absolute because climatic



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. 3. Biosolids composting mass balance diagram (15).

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 7–10 (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:

$$X_R = \frac{X_C\ (S_M\ -\ S_C)\ +\ X_A\ (S_M\ -\ S_A)}{(S_R\ -\ S_M)} \tag{5}$$

The ratio W is also recalculated:

$$W = \frac{X_C\ (1\ -\ S_C)\ +\ X_R\ (1\ -\ S_R)\ +\ X_A\ (1\ -\ S_A)}{X_C\ S_C\ V_C\ k_C\ +\ X_R\ S_R\ V_R\ k_R\ +\ X_A\ S_A\ V_A\ k_A} \tag{6}$$

If W is still not <10, 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, because it is not as important as the structural integrity of the pile. The recycle rate, X_R, and makeup supply are calculated using Eqs. (7) and (8).

$$X_R = (1 - f_2) f_1\ X_C \tag{7}$$

$$X_B = f_1\ X_C\ -\ X_R \tag{8}$$

where f_1 is defined as the ratio of external bulking agent (recycle and makeup) to biosolids:

$$f_1 = \frac{X_R\ +\ X_B}{X_C} \tag{9}$$

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:

$$f_2 = \frac{X_B}{X_B + X_R} \tag{10}$$

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 0.75–1.25, and for f_2 is 0.20–0.40. Once these values are chosen, the amount of recycled bulking agent (X_R) and new external bulking agent (X_B) can be calculated using Eqs. (7) and (8).

The value of the ratio W is then calculated using Eq. (2), indicating no amendment is used ($X_A = 0$). If W is ≤10, then the mixture has sufficient energy to compost. If W is >10, 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

Table 4
Densities of Compost and Bulking Agents

Material	Density(lb/yd^3)
Digested sludge	1500–1700
Raw sludge	1300–1700
New wood chips	445–560
Recycled wood chips	590–620
Finished compost	930–1040

Source: From ref. 15.
1 lb/yd^3 = 0.595 kg/m^3.

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 Eq. (2). Increasing the amount of amendment until W is <10 will result in the proper compost energy balance. 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 7 to 10 at this operation (39).

The best means to determine 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 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, CA (15).

In the process biosolids are converted to a relatively stable organic residue, reduced in volume by 20–50%. 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 3–7 ft (1–2 m).

Based on processing biosolids containing 20% solids, land requirements for the windrow process are higher than for the aerated pile process. It has been estimated that an extra 25% land area is needed for the windrow process based on specification such as 5 ft (1.5 m) high and 7 ft (2 m) wide with a 2 wk 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 40–50%. 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 9–15:1. If wood chips are used as the bulking agent, the C/N ratio will be raised to approx 20–30:1 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 10 d. Temperatures in the outer layers are considerably cooler and may approach atmospheric conditions. During wet periods and winter, maximum temperatures may only be 130–140°F (55–60°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 113–122°F (45–50°C). These variations in temperature are illustrated in Fig. 4.

Large-scale, 270 dry t/d (243 T/d) processing of digested primary biosolids (23% 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

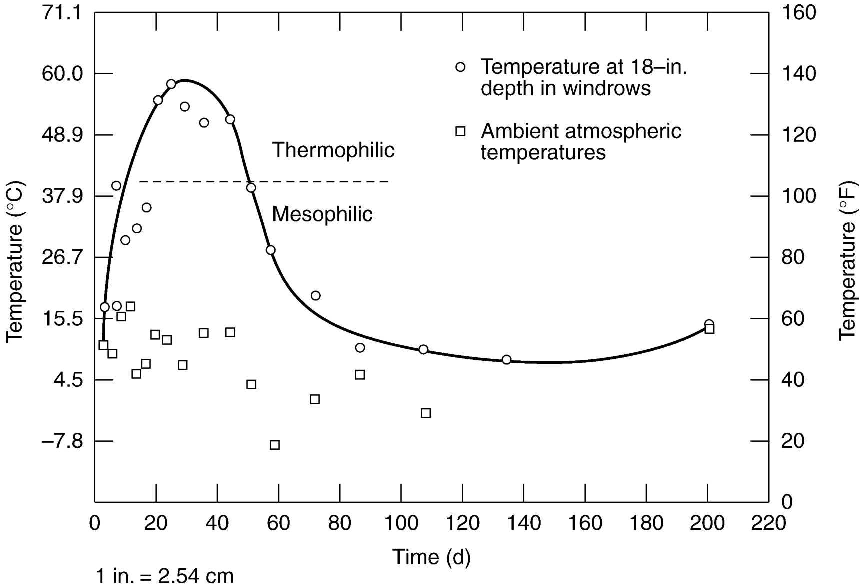


Fig. 4. Temperature profile of a typical compost windrow (15).

tests with windrows, undigested primary and waste activated sludge biosolids were found to produce offensive odor (4). Also, composting of digested biosolids did not kill all seeds, and these were present in the final product.

The Los Angeles County Sanitation Districts are currently composting digested, centrifuged primary biosolids (23% solids) in windrows mixed with recycled composted biosolids (60% solids) in 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 more times a day for the first 5 d to mix the material completely, minimize odor, and ensure sufficient oxygen transfer. The biosolids are then turned once 1 d for about 30 d, 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 3400 t/h (3084 T/h) of a density of 1890 lb/yd^3 (1120 kg/m^3). This is equivalent to a volume capacity of 3600 yd^3/h (2752 m^3/h). Three machines operate continuously for two shifts 1 d. 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 in. (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 in. (0.38 m) deep and 15 ft (4.5 m) wide was placed on the paved area. Biosolids (20–25% solids) were distributed to the chips at 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 1 wk, flattened after 2 wk to a 12 in. (0.30 m) layer and harrowed for further drying, generally to more than 65% solids. The material was then removed from the windrow area and stockpiled for an additional 30 d 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 25–30% lost in the composting process. Some of the bulking agent was consumed in the biological oxidation processes during composting, and a large portion was loss in the screening process.

6.2. Energy Requirements

Thermodynamic considerations in the composting of biosolids are discussed in an article by Haug and Haug (39). As indicated previously, the reaction is self-sustaining when the ratio W is <10. More 80% 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 10–50 dry t/d (9–45 t/d) operation been estimated (41):

a. Labor: 1.8–3 h.
b. Gasoline: 1.1 gal (4.5 L).
c. Diesel Fuel: 3.3–4 gal (13.5–16.5 L).
d. Electricity: 3–8 kWh (12–32 MJ).

Where finished compost is used as the bulking agent, and increased windrow turning frequency is practised, 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 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 10 d of composting in the interior of windrows. For interior samples, final compost coliform concentrations of less than 1/g 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.

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.

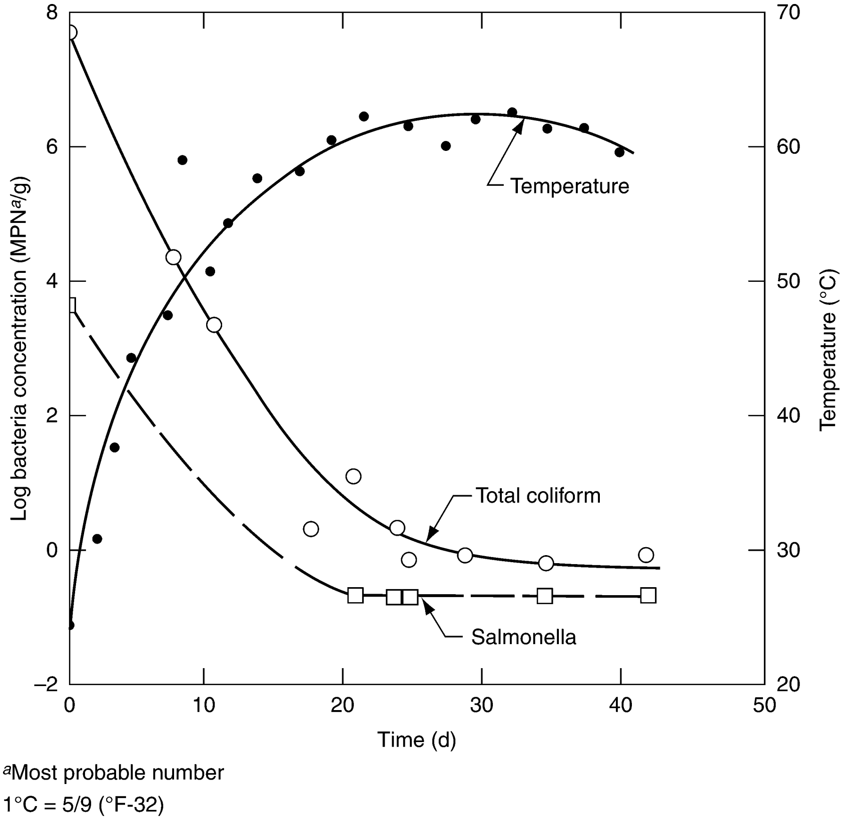


Fig. 5. Destruction of pathogenic organisms as a function of time and temperature during the composting of undigested biosolids by the windrow method (15).

Interruption of regular turning of the biosolids may cause odor problems, because compost windrows quickly become anaerobic under these circumstances. Unpleasant odor 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 practised 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 period. Although recent experiments have shown high concentrations of the fungus *A. 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* (46–50).

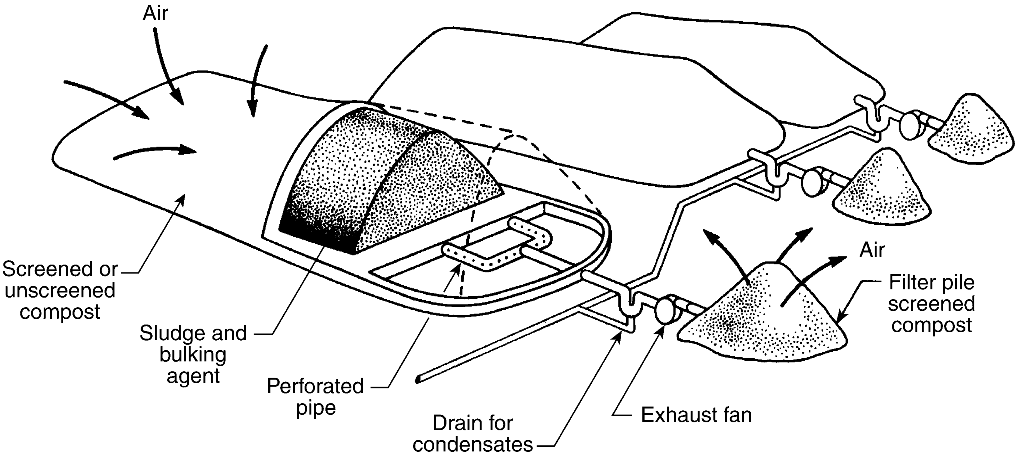


Fig. 6. Configuration of individual aerated piles (15).

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 Fig. 6.

Wastewater biosolids are converted to compost in approx 8 wk 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% 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 directed into a small pile of screened, cured compost where odorous gases are effectively absorbed. After about 21 d 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.
c. **Drying and Screening.** Drying of 40–45% 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 in. 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 d to assure no offensive odor 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). Because composting times tend to be slightly shorter and anaerobic conditions can be more readily prevented, the risk of odor is reduced. Two distinct aerated static pile methods have been developed, the individual aerated pile and the extended aerated pile.

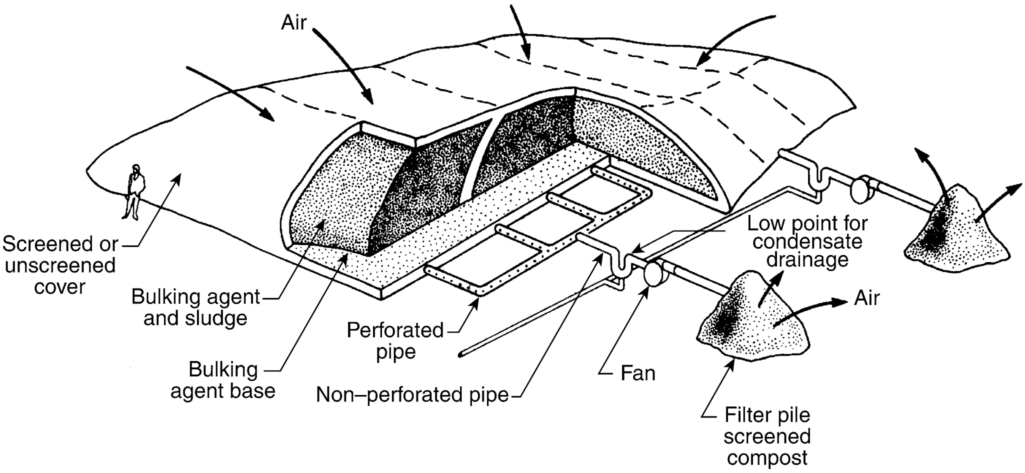


Fig. 7. Aeration pipe set-up for individual aerated pile (15).

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, 4–6 in. (10–15 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 is terminated at least 8–10 ft (2–3 m) inside the ends of the pile. A nonperforated pipe that extends beyond the pile base is used to connect the loop of perforated pipe to the blower (*see* Fig. 7).

About 6-8 in. (15–20 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% solids and two volumes of bulking agent. The resulting mixture contains 40% 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-in. (0.3 m) layer of cured, screened compost or an 18 in. (0.4 m) layer of unscreened compost. This outer blanket of compost provides insulation and prevents escape of odor during composting. Unstabilized biosolids can generate odor during dumping and initial pile construction. However conditioning with lime during dewatering will minimize this. The nonperforated 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 min on and 15 min off for a 56 ft (17 m) long pile containing up to 80 wet t (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 approx 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 more than 50%. The odor filter pile should contain 1 yd^3 (0.76 m^3) of screened compost for each 4 dry t (3.6 T) of biosolids in the compost pile. Filter piles are sometimes constructed with a 4-in. (10 cm) base layer of wood chips to prevent high back pressures on the blower.

Land area requirements are estimated at 1 acre/3–5 dry t (1 ha/6.7–11.2 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 1 acre/5 dry t (1 ha/l1.2 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 1 d 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 d. The first section is removed after 21 d. 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 8 shows such a system. The area requirement of an extended pile system is about 50% 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%.

The aerated pile system has proven effective on a full-scale basis at Beltsville, MD; 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 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, more 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 2–3 d to dry off the moisture. At other locations, the airflow 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% 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 airflow

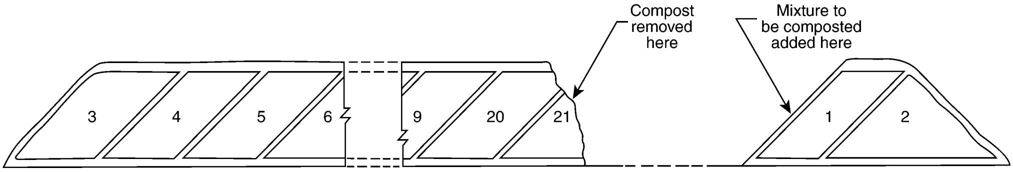


Fig. 8. Configuration of extended aerated pile (15).

must be reduced. Aeration cycles of 20–30 min with the fan operating 1/10–1/2 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 condensed moisture 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 odor if allowed to accumulate in puddles around the piles. Data is not available on combined leachate and condensed water characteristics; the quantity may, however, vary from 6 to 20 gal/d (22–75 L/d) per pile containing 50 yds^3 (38 m^3) of biosolids during dry weather (53).

7.5. Bulking Agents

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%, 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% results; this is well in excess of the 40–50% volatility achieved in the mixture of digested biosolids and recycled compost. Therefore, volatility content is 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 10–50 dry t/d (9–45 t/d) operation (20% biosolids) is listed as follows (41):

a. Labor: 1.5–2.8 h.
b. Wood chips: 2 –8 yd^3 (2.1 m^3).
c. Gasoline: 1.1 gal (4.1 L).

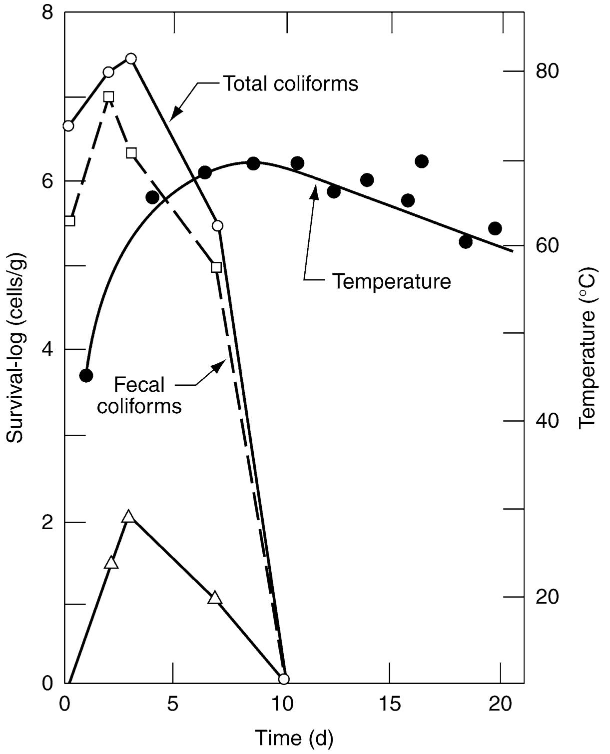


Fig. 9. Destruction of pathogenic organisms as a function of time and temperature during composting of undigested biosolids by the aerated pile method (15).

d. Diesel Fuel: 2.7–3.5 gal (10.2–13.2 L).
e. Electricity: 7.5–17.5 kWh (29.7–69.3 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 10th d of composting. Studies using "F" bacteriophage and virus as an indicator showed that the virus was essentially destroyed by the 13th d. 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 d 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 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

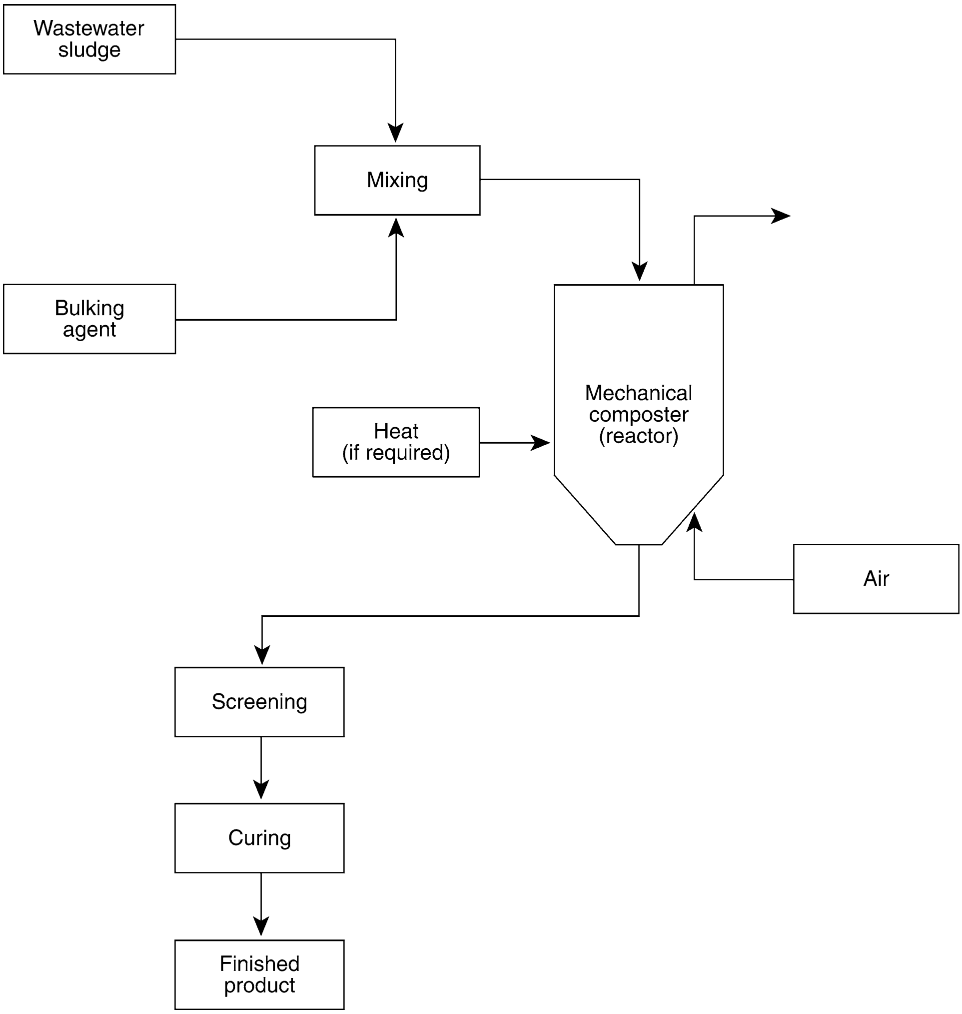


Fig. 10. Process flow diagram for confined composting system (15).

transport of aeration air between blower and odor filter pile. Moisture content within odor filter piles should be kept less than 50%. 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 odor 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 Fig. 10.

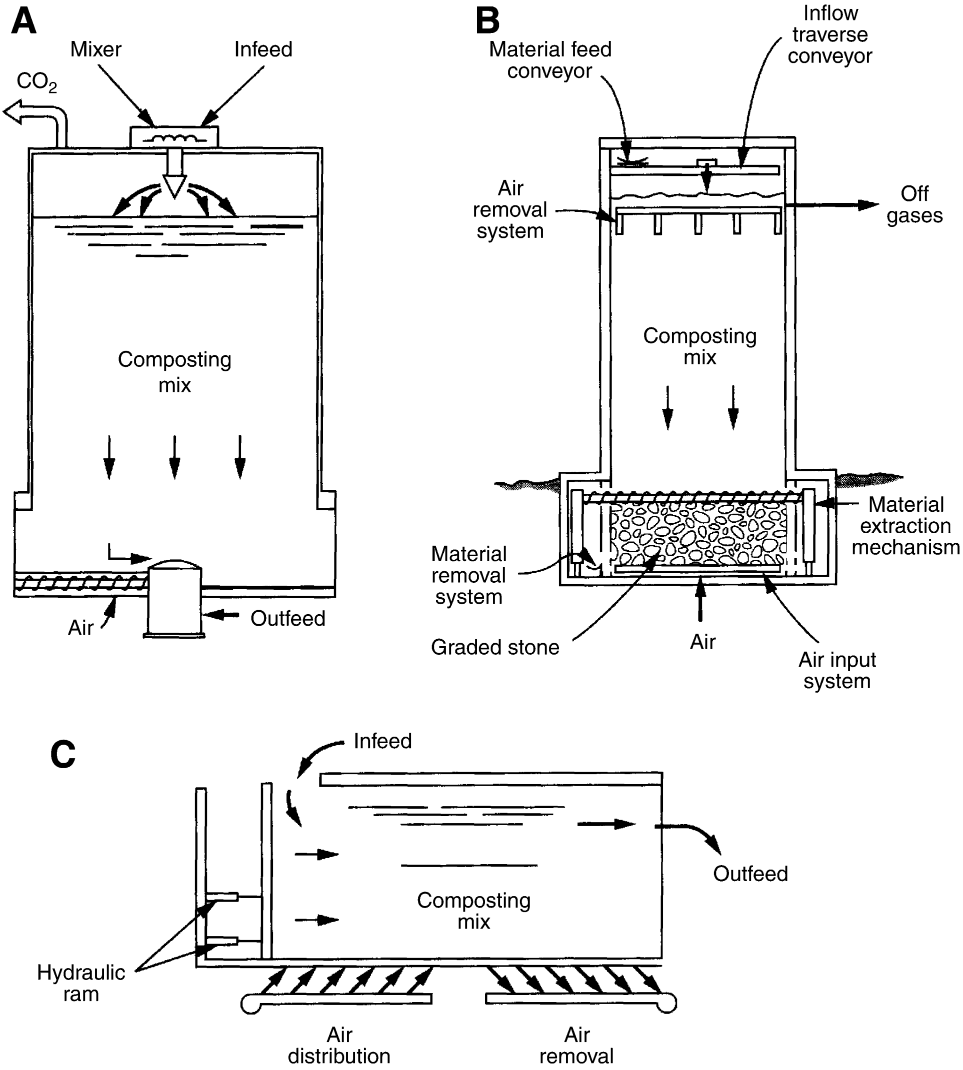


Fig. 11. Plug-flow in-vessel composting bioreactors (66): **(A)** cylindrical; **(B)** rectangular; **(C)** tunnel.

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 point (1). Generally air is 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 Fig. 11, and agitated bins shown in Fig. 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

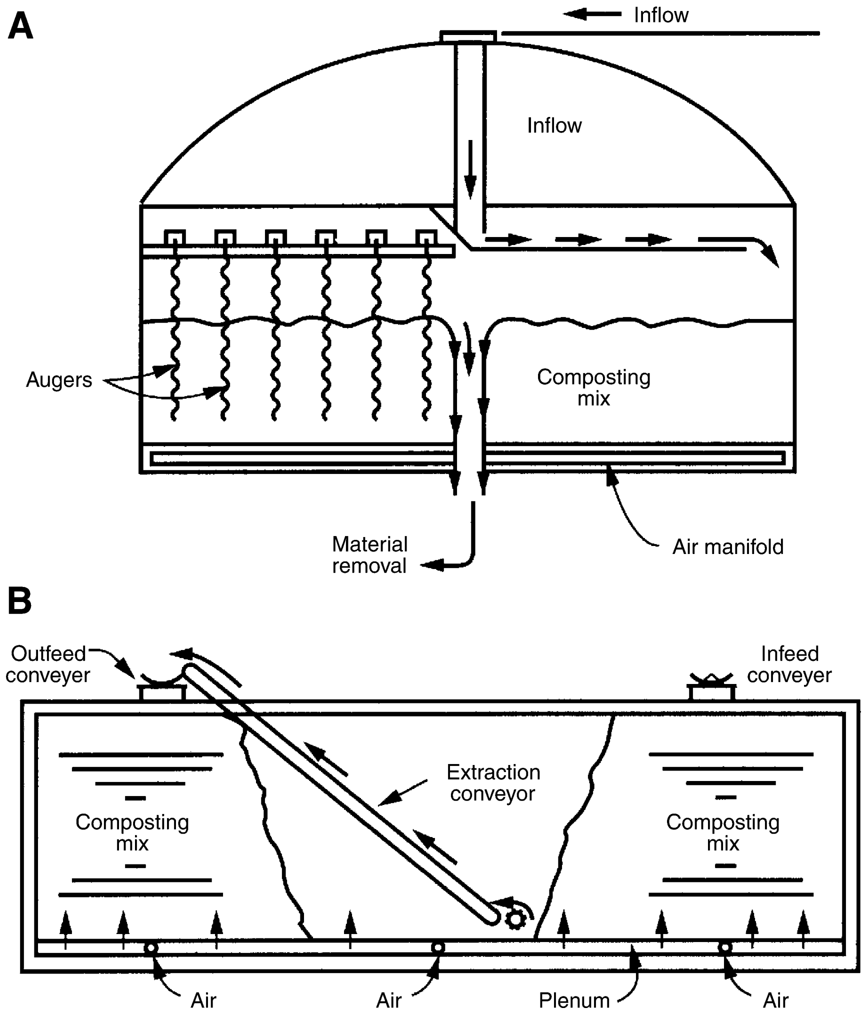


Fig. 12. Agitated (mixed) in-vessel composting bioreactors (66): **(A)** circular; **(B)** rectangular.

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 Fig. 13.

In vertical plug-flow systems, the biosolids and bulking agent mixture are introduced into the top of the reactor vessel and compost is discharged out of 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 exhausted through slots in the floor of the reactor. Compost is discharged from the end of the reactor opposite the ram.

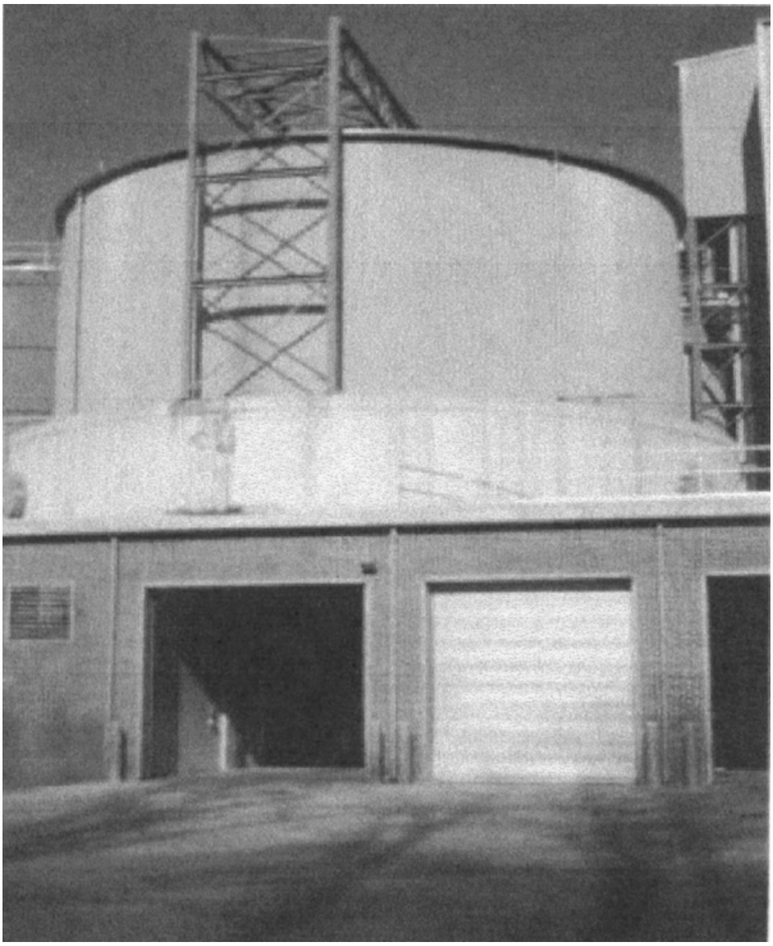


Fig. 13. Cylindrical composting vessel (1).

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% of both capital and operation and maintenance costs (18). Composting facilities usually use either wet scrubbers or biofilters for odor control. 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 10–21 d 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 d instead of the 20 d used in the unconfined systems, are usually specified by in-vessel equipment manufacturers (59). In-vessel technology offers the following advantages (11,18):

a. The composting process can be more closely controlled.
b. The effects of weather are diminished.
c. Less bulking agent may be required.
d. The quality of the resulting product is more consistent.
e. Less manpower is required to operate the system and staff is less exposed to the composting material.
f. Process air can be more easily collected for treatment to reduce odor emissions.
g. Less land area is required.
h. 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. Generally 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 odor 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 odor (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:

a. Bulking agents must be added.
b. Turning, monitoring, or process control is necessary.
c. Feed and finished materials must be moved with mechanical equipment.
d. Storage piles must be maintained for curing and distribution.
e. 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

Table 5
Comparison of Composting Methods

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

Source: From ref. 1.

their application to cases in which the ultimate use of the compost is firmly established to justify the investment. Table 5 compares the three composting methods and highlights key features of each. Detailed information on composting can be found in refs. 60–69.

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 odor. 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 40,700 to 103,300 USD/dry T (37,500–92,800 USD/dry t)/d processing capacity. A typical aerated static pile facility costs approx 40,700 USD/dry T (37,500 USD/dry t)/d of processing capacity (70,71).

Typical operation and maintenance (O&M) costs for in-vessel systems range from 188 to >250 USD/dry t/d. Aerated static pile O&M costs average 188/dry t/d (70,71). Costs for windrow systems fall between the costs for in-vessel and aerated static pile. The selling price for compost ranges from 6 to 12 USD/yd^3 or 12–25 USD/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 2006 costs using the Cost Index for Utilities (Appendix); all costs were multiplied by a factor of 528.12/424.91 = 1.24 (72).

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% solids) is generated at the rate of 1 dry t/MG (0.00024 T/m^3). The compost facility will handle 10 dry t/d (9 T/d) at 20% solids, 7 d/wk. 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. However, this design example 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 approx 40–50% and volatile solids content will be reduced by about 20–40%. The density will decrease by 15–25% because of evaporation.
b. The values for the process variables are assumed to be as follows:

$S_C = 0.20$ $S_R = 0.70$ $S_A = 0.90$ $S_M = 0.40$
$V_C = 0.50$ $V_R = 0.35$ $V_A = 0.90$ $V_M = 0.50$
$k_C = 0.45$ $k_R = 0.15$ $k_A = 0.50$

c. If the mixture has a high ratio of water to degradable organics by weight (W ratio is >10), amendment will be added to reduce W.

Solution

The amount of finished compost to be recycled can be calculated using Eq. (3).

$$X_R = \frac{X_C\,(S_M - S_C)}{(S_R - S_M)} \tag{3}$$

$$X_R = \frac{50\,(0.04 - 0.20)}{(0.70 - 0.40)}$$

$$X_R = 33.3 \text{ t/d } (30.3 \text{ T/d})$$

This indicates that if a mixture moisture content of 40% 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 Eq. (4) in order to determine whether to compost.

$$W = \frac{X_C\ (1 - S_C) + X_R\ (1 - S_R)}{X_C\ S_C\ V_C\ k_C + X_R\ S_R\ V_R\ k_R} \quad (4)$$

$$W = \frac{50(1 - 0.20) + 33.3(1 - 0.70)}{50(0.20)\ (0.50)\ (0.45) + 33.3\ (0.70)\ (0.35)\ (0.15)}$$

$$W = 14.4$$

The calculated value for *W* is too high, indicating that amendment addition is required. Increasing the recycle rate to create a mixture of 50% 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 t (0.9 T) amendment per 10 t (9 T) of biosolids cake is added to the mixture, the recycle rate can be calculated using Eq. (5):

$$X_R = \frac{X_C\ (S_M - S_C) + X_A\ (S_M - S_A)}{(S_R - S_M)} \quad (5)$$

$$X_R = \frac{50\ (0.40 - 0.20) + 5\ (0.40 - 0.90)}{(0.70 - 0.40)}$$

$$X_R = 25.0 \text{ t/d } (22.7 \text{ T/d})$$

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 Eq. (6):

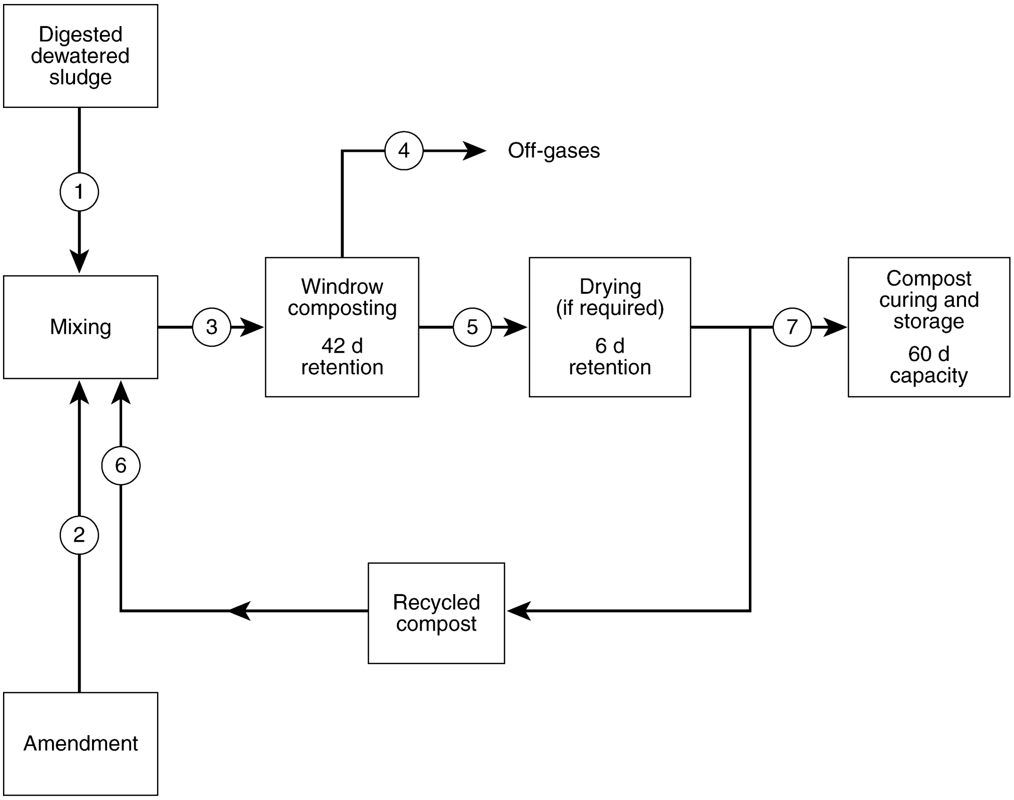
$$W = \frac{X_C\ (1 - S_C) + X_R\ (1 - S_R) + X_A\ (1 - S_A)}{X_C\ S_C\ V_C\ k_C + X_R\ S_R\ V_R\ k_R + X_A\ S_A\ V_A\ k_A} \quad (6)$$

$$W = \frac{50(1 - 0.20) + 25(1 - 0.70) + 5(1 - 0.90)}{50(0.20)\ (0.50)\ (0.45) + 25(0.70)\ (0.35)\ (0.15) + 5(0.90)\ (0.90)\ (0.50)}$$

$$W = 9.2$$

This mixture of biosolids cake, recycled compost, and amendment is self-sustaining and will degrade properly. Figure 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 6 wk of turning is approx 65% 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 5 d to mix the materials completely, to minimize odor, and to ensure sufficient oxygen transfer. After the initial 5 d 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/yd^3)	Volume (yd^3)	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

1 t = 0.907 Tonne
1 lb/yd^3 = 0.6 kg/m^3
1 yd^3 = 0.76 m^3

Fig. 14. Process flow diagram-windrow composting of biosolids from 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:

a. Minimize hauling and handling cost.
b. Maximize use of existing equipment in the compost operation.
c. Minimize the use of amendment which adds to the cost and is not recoverable.

Table 6
Densities of Compost and Bulking Agents Used in Example 2

	Bulk density	
Constituent	lb/yd^3	kg/m^3
Dewatered sludge (20% solids)	1600	960
New wood chips	500	300
Recycled wood chips	600	360
Screened compost	865	519
Unscreened compost	1000	600

Source: From ref. 15.

d. 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 as follows:

a. The bulking agent to biosolids ratio.
b. The ratio of new to recycled bulking agent.

The materials balance in this example is based on the following assumptions:

a. Biosolids to be composted are 50 wet t/d (45 T/d) of undigested biosolids, 7 d/wk, with no digestion.
b. 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.
c. Three-fourths of the chips are recovered by screening and reuse.
d. The water content and total weight of the compost mixture is reduced by approx 30–40% and volatile solid's content is reduced by about 10–15%. The density decreases 15–20% because of evaporation.
e. The extended aerated pile system will be used.

Information on the bulk density of biosolids is surprisingly scarce. Tests conducted at Beltsville 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 6:
It is also assumed that the process variables have the following values:

$S_C = 0.20$ $S_B = 0.70$ $S_R = 0.70$
$V_C = 0.75$ $V_B = 0.90$ $V_R = 0.80$
$k_C = 0.45$ $k_B = 0.10$ $k_R = 0.10$

Solution

a. **The bulking agent to biosolids ratio.** Biosolids composting will operate 5 d/wk, 8 h/d using the extended aerated static pile method. The volume to be composted per work day is as follows:

$$(50 \text{ wet t/wk-d}) \ (7 \text{ wk-d/5 work-d}) = 70 \text{ wet t/work-d } (63.5 \text{ t/work-d})$$

It is assumed that the dewatered biosolids arrive onsite 5 d/wk from the dewatering operation which runs only 5 d/wk. Equalization storage to cover weekend operation of the lant 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 Eqs. (7) and (8) and assuming:

$$f_1 = 0.75 \text{ and } f_2 = 0.25$$

$$X_R = (1 - f_2) f_1 X_C \tag{7}$$

$$X_B = f_1 X_C - X_R \tag{8}$$

$$X_R = (1 - 0.25)\ (0.75)\ (70) = 39.4 \text{ t/d } (35.7 \text{ T/d})$$

$$X_B = (0.75)\ (70) - 39.4 = 13.1 \text{ t/d } (11.9 \text{ T/d})$$

The ratio W can be calculated using Eq. (2):

$$W = \frac{X_C\ (1 - S_C) + X_A\ (1 - S_A) + X_B\ (1 - S_B) + X_R\ (1 - S_R)}{X_C\ S_C\ V_C\ k_C + X_A\ S_A\ V_A\ k_A + X_B\ S_B\ V_B\ k_B + X_R\ S_R\ V_R\ k_R} \tag{2}$$

$$W = \frac{70\ (1 - 0.2) + 39.4\ (1 - 0.7) + 13.1\ (1 - 0.7)}{70\ (0.2)\ (0.75)\ (0.45) + 39.4\ (0.7)\ (0.9)\ (0.1) + 13.1\ (0.7)\ (0.8)\ (0.1)}$$

$$W = 9$$

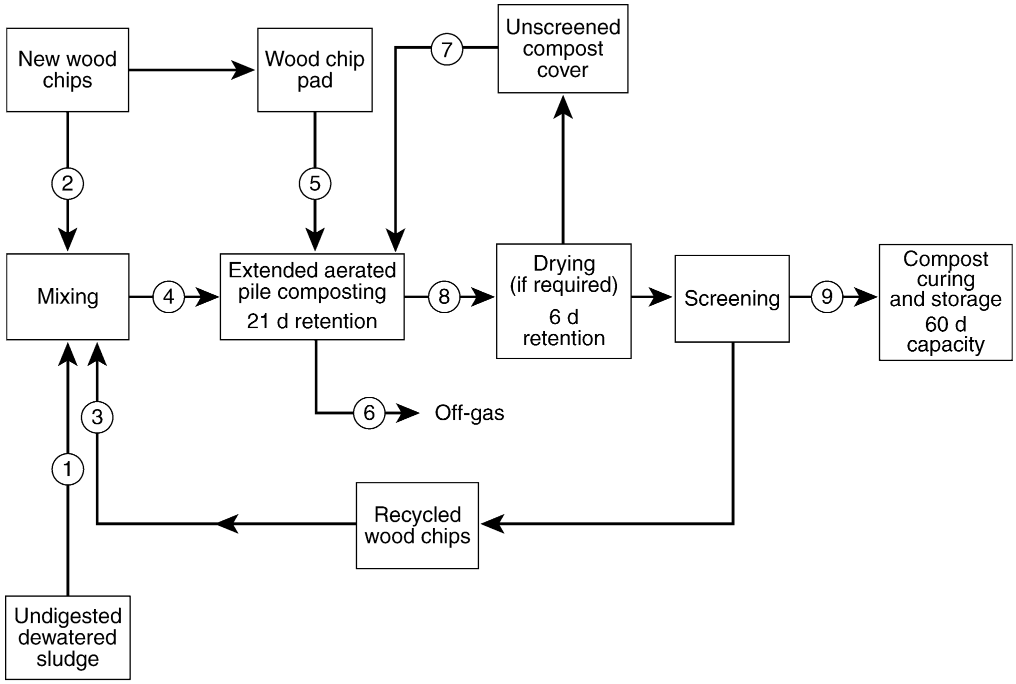
Because $W<10$, no amendment addition is required.

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:

$$(50 \text{ ft})\ (18.5)\ (1 \text{ ft})/27 \text{ ft}^3/\text{yd}^3 = 34.3 \text{ yd}^3/\text{d } (26.2 \text{ m}^3/\text{d})$$



5 (d/wk) operation

Location	Wet (t)	Percent solids	Dry (t)	Density (lb/yd^3)	Volume (yd^3)	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	90
4	122.5	41	50.8	900	271.2	90
5	6.6	70	8	500	34.3	90
6	58.7	–	10.9	–	–	–
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

1 t = 0.907 T
1 lb/yd^3 = 0.6 kg/m^3
1 yd^3 = 0.76 m^3

Fig. 15. Process flow diagram for the extended pile compost facility for 10 MGD-activated sludge plant (15).

Unscreened compost is required each day to cover the pile. This layer will be 18 in. (0.46 m) thick:

$$(50\ \text{ft})\ (18.5\ \text{ft})\ (1.5\ \text{ft})/27\ \text{ft}^3/\text{yd}^3 = 51.4\ \text{yd}^3/\text{d}\ (39\ \text{m}^3/\text{d})$$

Figure 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 in. (10-cm) diameter perforated aeration pipe, 50 ft (15 m) of nonperforated pipe, three 4 in. (10-cm) tee connectors, and one blower/timer unit with weather protection and condensed 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 t (1.3 L/s/T) of biosolids in the daily pile. Nonperforated pipe should be used to connect the aeration pipe loop to the blower. The exhausted air will be filtered in a

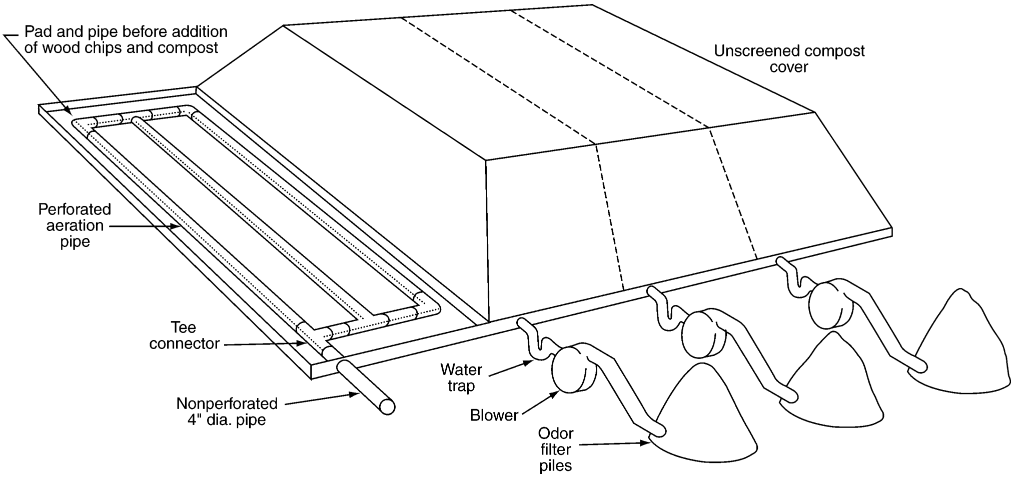


Fig. 16. Design example-extended aerated pile construction (15).

pile of screened compost. The filter pile will contain at least one yd^3 of material/ 30 wet t (1 m^3/35.5 T) of biosolids in the daily pile or 4 yd^3 (3 m^3) for this design. Fig. 16 illustrates this design example. The minimal area requirements for various composting site components are shown in Table 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 <40% or >50%.

Approximately 8.5 ft^3 of air/min/t of dry solids (4 L/s/T) in the pile is required. At Beltsville, this was delivered by a centrifugal fan operating at 5 in. differential water pressure (1.25 kN/m^2) (41). The Bangor, Main system uses a 1/3 hp (0.25 kW) blower rated at 335 ft^3/min (158 L/s) at 5 in. 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 5–15% 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 many individual blowers would be eliminated. On the other hand, a central blower system is not especially flexible. Because it is important to maintain the proper aeration rates in each pile, an airflow 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

Table 7
Minimal Composting Area Requirements for Example 2[a]

Function	Area required ft^2	Area required m^2
Truck unloading and mixing	5000	465
Composting (28 d) (50) (18.5) (1.15 excess)	30,000	2792
Unscreened compost	10,000	931
Drying and screening	20,000	1862
Compost curing and storage (60 d) (200 yd^3/d) (27 wet t) (10 ft deep) + excess	33,000	3071
New wood chip storage (60 d) (87 yd^3/d) (27 wet t) (12 ft deep) + excess	15,000	1396
Subtotal	113,000	10,517
Maintenance building, operations building and laboratory, lunch room, and locker room	4000	372
Employee and visitor parking	5000	465
Miscellaneous storage	1000	93
Subtotal	10,000	930
Total	123,000	11,447

Source: From ref. 15.
*Note:*123,000 ft^2 (11,447 m^2) = 3 acres (1.14 ha).
Land utilization = 6.6 dry t/acre (14.8 T/ha).
[a] 50 wet t/d (45 T/d) 10 dry t/d (9 T/d)

control. The disadvantages of this type of combined system are the high initial cost and the reduced flexibility of operation. Possible elimination of the 1 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 per day
X_A	Total wet weight of amendment per day
X_R	Total wet weight of recycle per day
X_B	Total wet weight of external bulking agent per day

X_M	Total wet weight of mixture per day
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) for digested; (0.60–0.80) for 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 (0.00–0.90)
V_B	Volatile solids content of 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 (0.00–0.20)
k_B	Fraction of external bulking agent volatile solids degradable under composting conditions (0.00–0.40)
k_M	Fraction of mixture volatile solids degradable under composting conditions (0.20–0.60).
t	Ton (English)
T	Tonne (Metric)

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APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers[a]

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

Source: [a]From ref. 72.

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Vermicomposting Process

Lawrence K. Wang, Yung-Tse Hung, and Kathleen Hung Li

CONTENTS

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 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 as fish bait or animal protein supplement. Earthworm marketing is a complex problem; for municipal biosolids applications, surplus earthworms might be considered as a byproduct, 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.
f. Future development and directions of the process.

From: *Handbook of Environmental Engineering, Volume 6: Biosolids Treatment Processes*
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1.2. Process Description

Vermicomposting differs from the conventional composting of wastewater treatment of 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 (1–66). 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.

Figure 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, in which raw liquid sludge is placed in beds. If anaerobic digestion is used before earthworm conversion, additional pretreatment might be required. Especially if moisture is high, a bulking agent such as wood chips may be useful in some cases for keeping the bed porous and aerobic. However, biosolids are 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 design and 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. (1) of the State University of New York at Syracuse, College of Environmental Science and Forestry in 1977. Later, Mitchell et al. (2) investigated the potential role of the earthworm (*Eisenia foetida*), on the decomposition of sewage biosolids in drying beds and reported the results in 1980. Specifically, Mitchell et al. sought to determine the 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 was 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, 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 cost for a municipal operation serving (a) a community of 50,000 persons and (b) a community of about 500,000 persons. Vermicomposting was compared with three other methods of solid waste management: sanitary landfill, windrow composting, and combustion.

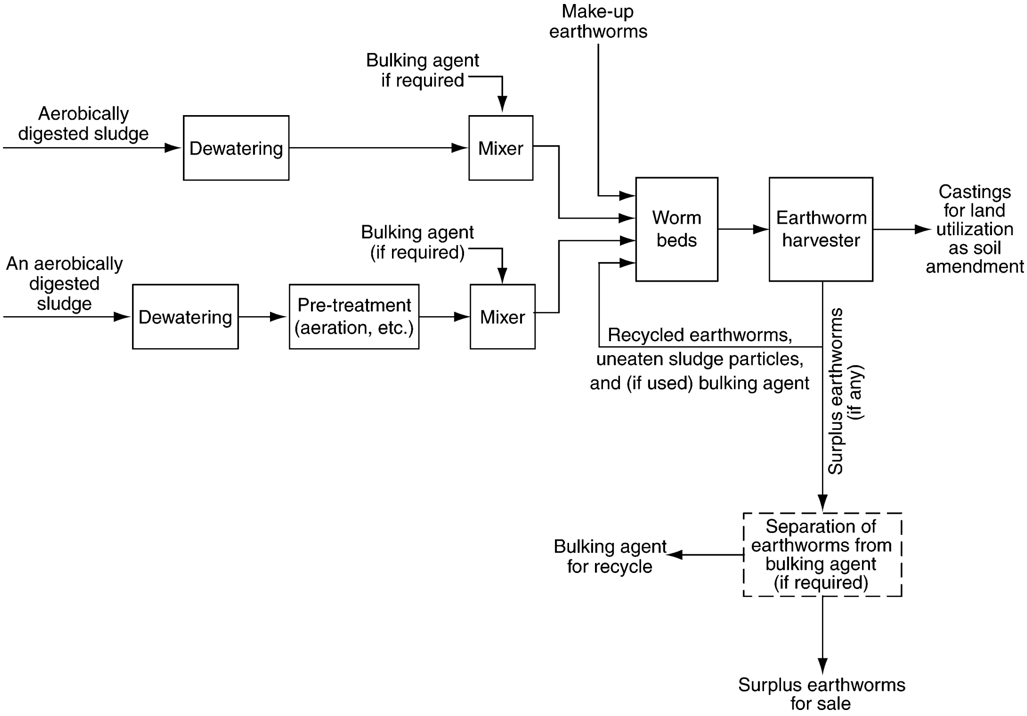

Fig. 1. Diagram of an earthworm process.

In 1980, vermicomposting was estimated to cost about 24–32 USD/t of waste processed. (*Note*: 1 t = 2000 pounds; 1 pound = 0.454 kg.)

In 1981, Hornor and Mitchell (4) studied the effect of the earthworm, *E. 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 biosolids was examined in detail. Results of tests performed by Hartenstein on two earthworm species—*Eudrilus 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.
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, 90 t of earthworm manure was produced from the biosolids more than a 5-yr period. Different size of 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 4:1. Their 1981 cost analysis showed the system to be cost effective at a cost of 29.45 USD/dry t in a 10 t/d facility, and to return a profit of 3.34 USD/dry t if castings were produced at the rate of 50 t/d.

In 1982, Hartenstein (8) reported: (a) the metabolic parameters of the earthworm *E. 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*, *Pheretima hawayana,* and *Perionyx 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%. The best growth and cocoon production for this earthworm species was shown to occur at temperatures of 20–25°C. With both dewatered and liquid biosolids, vermistabilization units functioned successfully for long periods of time—up to 1 yr for dewatered biosolids and at least 6 mo 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. (12) of Cornell University 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*, *E. foetida*, *E. eugeniae*, *P. excavatus*, and *P. hawayana* were used in one or more of the studies. The best growth and reproduction of these species occurred at temperatures of 20–25°C. Growth of all five species was reduced at 30°C and death occurred at 35°C. Of the five species, *E. foetida* produced the largest number of young in a 20 wk study. The growth of *E. foetida* occurred optimally in media with a total solids content, wet basis, of between 9 and 16%. 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. (13) of the University of Texas at Austin, TX. 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 140–198 d 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 waste-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 4–20%. Loading rates of about 21,000 g/m^2/wk volatile solids or less resulted in satisfactory operation of LSVS reactors stabilizing liquid primary and liquid waste-activated sludge. Loading rates greater than 1200 g/m^2/wk 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%, a considerable increase from the 0.6 to 1.3% 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 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, 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, along with 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 d. 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 United States 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 *E. foetida* is known to contain bactericidal enzymes. In 1990, Amaravadi et al. (18) tested the earthworm for virucidal activity using Cowpea mosaic virus and Tobacco mosaic virus as model agents. 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 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 United States Patent no. 5055402, entitled, "Removal of Metal Ions With Immobilized Metal Ion-Binding Microorganisms" was awarded to Greene et al. (19) on October 8, 1991. 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 United States 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, which were divided into two general categories. First, earthworms were directly used in horticulture and agriculture to enhance crop growth. Second, earthworms were used to turn various residuals into beneficial composts for reuse. However, despite the increasing number of studies, financial support for vermicomposting research has been cut by the funding agencies in the United States since 1990.

Another problem is the process's failure to meet regulatory requirements. The United States Environmental Protection Agency's (US EPA) "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 3 d. Worms can survive in thermophilic composting windrows, but they tend to stick to the edges of the pile. Temperatures more than 35°C is the heat generated by thermophilic composting, which is too high for earthworms, and will kill them. In vermicomposting, temperatures are generally kept less than 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 owing to the aforementioned above technical and legal problems.

There has been continuous debate in the State of California, regarding the classification and potential regulation of composting facilities. A draft of regulations released in August 1994 by the California Integrated Waste Management Board 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 US 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 6 mo. Whereas with the modified vermicomposting which includes thermophilic composting pretreatment, and vermicomposting post-treatment, it is possible to create a marketable end product in one-sixth of the operating time. Compared with 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 had paid 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) has developed a proprietary "Vermiconversion System," which significantly modifies traditional vermiculture windrow methods. Variations include sloped plastic liner beneath the windrow, reclaimed 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 m^2 = 0.4046 ha) facility currently processing around 100 t/d 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 they are." Leafy greens, wood fines, and manures proceed through a blending plant, then "cured" through 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. Continuously 2–4 in. of material are applied every other day. The rows are compartmentalized to prevent possible contamination of the entire facility. The facility adopts both the thermophilic composting pretreatment (for 3–15 d aiming at pathogen reduction and decomposition) and the vermicomposting post-treatment (for additional 15–30 d aiming at final curing and decomposition). Their worm castings product is being sold for 33 USD/t on the bulk market. The company is now building a 100-acre facility to manage San Diego's biosolids under a 20 yr 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 ft trough (*Note:* 1 ft = 0.3048 m) that is 2.5 ft deep and 8 ft wide, with a mesh floor. An adapted manure spreader makes a daily pass over the trough, laying down about three inches (*Note:* 1 in. = 2.54 cm) of prepared organic materials, or roughly 6 t/d (*Note:* 1 t = 2000 pounds; 1 pound = 0.454 kg). As the worms eat their way 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 2500 t 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 5 or 6 t of food scraps, more than 2 t of supplemental yard trimmings or compost, and around 0.5 t of paper per day. It takes only 21 d to make earthworm castings using the continuous flow system.

The Worm Concern (*Note:* It is "The Worm Connection" now in California) had grown to a 22-acre spread during its 18 yr in business. Around 100 t/d of brush, leaves, tree limbs, grass clippings, and horse manure is 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 (where 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 from one end and the unharmed worms come out from the other. 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) currently operate two sites, handling a combined total of 30 t/mo of composted yard trimmings from the Orange County landfill and 20 t/mo 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 United States

Since 1992 engineers and scientists in the countries other than the United States have shown their interest in the theories, principles, and applications of vermistabilization process. Practical applications of Vermicomposting process in disposal of biosolids and organic solid waste has been attempted by many entrepreneurs around the world. The progress in Vermicomposting process development and applications outside of the United States is discussed later (20–26).

In November 1992, Concheri et al. (20) of Italy reported humification of organic waste materials during earthworm composting. In March 1993, Anton et al. (21) of Spanish Council for Scientific Research, Madrid, Spain, reported carbofuran acute toxicity to *E. foetida* earthworms.

In 1993, Van-Gestel and Ma (22) of the National Institute of Public Health and Environmental Protection, Bilthoven, Netherlands, reported their results on development of QSAR's (quantitative structure activity relationship) in soil ecotoxicology. 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 8–10 t/yr from a bed area of 100 m^2 (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. (26) of the University of Vigo, Spain in 1995. 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 as 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 (7) 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." Collier and his wife still operate an earthworm farm in California, 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 of California at Santa Cruz, recently collaborated with the Colliers to complete his PhD research in a similar field.

Raymond C. Loehr of the University of Texas at Austin, Department of Civil Engineering, is another legend in vermistabilization technology development (11–13). If requested, he too consults on vermistabilization research and applications.

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 ecotechnology on all levels. Columns such as the following appear intermittently in each issue (27–29): Worm Shorts x New Products x International Worm News x The Industrious Worm (large-scale projects) x Hands-On x Worm Workers x Kids' Corner/Page x Questions and Answers x Eco-Logic x Worm Stories x Cyber-Worm x Advertisements and Resource Listings x 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 United States were terminated as a result of minor technical and legal problems, and a lack of financial and public support. However, it is encouraging to learn 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 earlier, there is even an annual International Symposium on Earthworm Ecology. Interest in vermistabilization process for biosolids management has quickly spread from the United States 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 co-operative 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 (40–44).

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 0.02–0.1 in. (0.5–3 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. If the oxygen uptake rate increases (46); the acid-extractable fraction of various nutrients also increases (50). The volatile content of the solids drops slightly and humic acid concentrations fluctuate (46). Although these effects may be beneficial, there is 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 is 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 later.

5.2. Process Operation and Troubleshooting

A number of potential operating difficulties and their solutions exist in the earthworm conversion process. However, none of these difficulties are insurmountable. 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:

a. *Worm drowning:* Worms must be protected from flooding.
b. *Worm loss as a result of migration from the process:* Caused by flooding, toxic sludge, unpalatable sludge, adjoining areas attractive to worms, lack of artificial lighting on rainy nights.
c. *Toxicity of sludge to worms:* Significant for anaerobically digested sludge. However, toxicity is eliminated by exposing the sludge to air for two months (48) or wetting sun-dried sludge daily for 14 d (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.
d. *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.
e. *Worm shortage in the process, so that worm additions are required:* Worms reproduce through 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).
f. *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. Therefore gradual start-up is desirable, especially for large plants. Also, earthworm exchanges may become available nationwide so that sludge operations can draw on larger numbers of earthworm suppliers.
g. *Temperature extremes:* Worm feed most rapidly at 15–20°C; at 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.

h. *Shortage of enzymes:* Not a problem, despite claims by marketers of enzyme preparations that these preparations are valuable to the process (52).
i. *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.
j. *Dehydration:* There is a minimum moisture content for the worm bed (52).
k. *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).
l. *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.
m. *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):

a. Earthworm conversion decreases the total nitrogen values in the biosolids, as ammonia nitrogen will be lost to the atmosphere.
b. Costs are unpredictable.
c. 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 2500 mg Cu/kg of wet substrate (55,56). Safe limits for these elements are not known.
d. Cadmium accumulates in the worm *E. foetida*. Zinc apparently does not accumulate in *E. 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.
e. Space requirements may rule out earthworm conversion at some treatment plants.
f. 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.
g. 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 *E. 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:

a. Detention time of sludge in worm beds = 2–32 d (49,50).
b. Worm reproductive cycle = 1–2 mo.

c. Rate of worm feeding (15°C) = 0.17–1.7 g dry sludge/g dry worm weight/d (48).
d. Optimum temperature = 15–20°C.
e. Dry matter content of worms = 20–25% (*E. foetida*) (49).
f. Minimum solids content of the worm bed mixture = 20%.

Actual minimum solids content depends on such factors as porosity, type of sludge, ability to maintain aerobic condition. Experiments are being conducted to better define these parameters.

6. PROCESS APPLICATION EXAMPLES

The Wright-Patterson Air Force Base in Dayton, Ohio (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 USD/d on transporting and disposing of waste. As the number of worms grow, so does the amount of waste they consume. The base acquired 250,000 worms in their climate-controlled home (at a constant 70°C 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 one 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, 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, 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 United States and other countries (66).

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23
Land Application of Biosolids

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

1. INTRODUCTION

Biosolids are essentially organic materials produced during wastewater treatment, which might be put to beneficial use. A popular example of such use is the addition of biosolids to soil to supply nutrients and replenish soil organic matter. Biosolids can be applied on agricultural land, forests, rangelands, or on disturbed land in need of reclamation (1). The thrust of recent legislation has been to encourage such beneficial recycling of biosolids through land application (2). The establishment of the industrial waste pretreatment programs (3) with the objective of reducing toxic pollutant loadings to municipal treatment facilities rendered more municipal biosolids suitable for reuse.

Wastewater biosolids may not always be used as a resource because of land acquisition constraints, the unavailability of agricultural land nearby or because they contain

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high levels of metals and other toxic substances. Because ocean disposal is no longer considered a viable or an appropriate alternative to utilization, land disposal through landfill has been optimized so that the increasing amount of biosolids generated by wastewater treatment plants can be accepted. Development of formalized methods for biosolids disposal to land is recent. Major efforts in this area have been funded by the United States Environmental Protection Agency (US EPA) since 1974 (4).

US EPA regulations (5) under Title 40 Code of Federal Regulations Part 503 (40 CFR 503) and their amendments (6), established the minimum national standards for the use and disposal of municipal biosolids. The reader is referred to refs. (7–11) for discussion and detailed information on the background, guidance, risk assessment, and applications of the regulations to control the recycling of biosolids through land application and disposal in landfills.

2. RECYCLING OF BIOSOLIDS THROUGH LAND APPLICATION

Recycling biosolids through land application serves several purposes. They improve soil properties, such as texture and water holding capacity, which make conditions more favorable for root growth and increase the drought tolerance of vegetation. Biosolids application also supplies nutrients essential for plant growth, including nitrogen and phosphorous, as well as some essential micro nutrients such as nickel, zinc, and copper (2). Biosolids can also serve as an alternative or substitute for expensive chemical fertilizers. The nutrients in the biosolids offer several advantages over those in inorganic fertilizers because they are organic and are released slowly to growing plants. These organic forms of nutrients are less water soluble and, therefore, less likely to leach into groundwater or run off into surface water (1).

Land application is well-suited for managing solids from any size wastewater treatment facility. As the method of choice for small facilities, it offers cost advantages, benefits to the environment, and value to the agricultural community. However, biosolids produced by many major metropolitan areas across the country are also land applied. For example, biosolids from the Blue Plains Wastewater Treatment Facility serving the District of Columbia and surrounding communities in Virginia and Maryland have been land applied since the plant began operation in 1930. The cities of Philadelphia, Chicago, Denver, New York, Seattle, and Los Angeles all land apply at least part of their biosolids production (1).

Land application is most easily implemented where agricultural land is available near the site of biosolids production, but advances in transportation have made land application viable even where hauling distance is more than 1000 miles. For example, Philadelphia hauls dewatered biosolids 250 miles to reclaim strip-mines in western Pennsylvania and New York City ships some of its biosolids more than 2000 miles to Texas and Colorado (1).

3. DESCRIPTION

There are several methods for land applications of biosolids. The selection of the method depends on the type of land and the consistency of the biosolids. Liquid biosolids are essentially 94–97% water with relatively low amounts of solids (3–6%). These can be injected into the soil or applied to the land surface. Specialized vehicles or modified tanker trucks are used to inject biosolids into the soil. These tankers have hoses leading from the storage tank to injection nozzles which release the biosolids. Biosolids applied to the land surface are usually incorporated into the soil with conventional farm equipment.

It is often economical to reduce the volume of biosolids before transportation or storage. The amount of water in biosolids can be reduced through mechanical processes such as draining, pressing, or centrifuging, resulting in a material composed of up to 30% dry solids (12–15). This material will be the consistency of damp soil. Dewatered biosolids do not require any specialized equipment and can be applied with conventional agricultural equipment, such as manure spreaders pulled by tractors.

The US EPA's 40 CFR Part 503, *Standards for the Use and Disposal of Sewage Sludge* (5), requires that wastewater solids be processed before they are land applied. This processing is referred to as stabilization and helps minimize odor generation, destroys pathogens (disease causing organisms), and reduces vector attraction potential. There are several methods to stabilize wastewater solids, including (12,16):

a. Adjustment of pH, lime or alkaline stabilization.
b. Anaerobic digestion.
c. Aerobic digestion.
d. Composting.
e. Heat drying.

The Part 503 Rule (5) defines two types of biosolids with respect to pathogen reduction, Class A and Class B, depending on the degree of treatment the solids have received. Both types are safe for land application, but additional requirements are imposed on class B materials. These are detailed in the Part 503 Rule and include such things as restricting public access to the application site, limiting livestock grazing, and controlling crop harvesting schedule. Class A biosolids (biosolids treated so that there are no detectable pathogens) are not subject to these restrictions.

In addition to stabilization, the Part 503 Rule sets maximum concentrations of metals which cannot be exceeded in biosolids that will be land applied. These are termed Ceiling Concentrations. Part 503 also establishes Cumulative Pollutant Loading Rates for eight metals which may not be exceeded at land application sites. A third set of metals criteria is also included in Part 503, known as Pollutant Concentrations. If these concentrations are not exceeded in the biosolids to be land applied, the Cumulative Pollutant Loading Rates do not need to be tracked. Table 1 shows the three sets of federal limits applicable to biosolids to be land applied (5,17).

The term *exceptional quality* is often used to describe a biosolids product which meets class A pathogen reduction requirements, the most stringent metals limits (Pollutant Concentrations), and vector attraction reduction standards specified in the Part 503 Rule. Vectors (flies, mosquitoes, rodents, birds, and so on) can transmit diseases directly to human or play a specific role in the life cycle of a pathogen as a host. Vector attraction reduction refers to processing which makes the biosolids less attractive to vectors thereby reducing the potential for transmitting diseases. Exceptional Quality biosolids products are as safe as other agricultural and horticultural products and may be used without site restrictions.

Table 1
Maximum Metal Concentrations

Metal	Ceiling concentration (mg/kg)	Cumulative pollutant loading rates (kg/ha)	Pollutant concentrations (mg/kg)
Arsenic	75	41	41
Cadmium	85	39	39
Copper	4300	1500	1500
Lead	840	300	300
Mercury	57	17	17
Molybdenum	75	NL	NL
Nickel	420	420	420
Selenium	100	100	100
Zinc	7500	2800	2800

Source: US EPA.
NL, no limit.

4. ADVANTAGES AND DISADVANTAGES

Land application offers several advantages as well as some disadvantages that must be considered before selecting this option for managing biosolids (1). Land application is an excellent way to recycle wastewater solids as long as the material is quality controlled. It returns valuable nutrients to the soil and enhances conditions for vegetative growth. Land application is a relatively inexpensive option and capital investments are generally lower than other biosolids management technologies. Contractors can provide the necessary hauling and land application equipment. In addition, on-site spatial needs can be relatively minor depending on the method of stabilization selected.

Although land application requires relatively less capital, the process can be labor intensive. Even if contractors are used for application, management oversight is essential for program success. Land application is also limited to certain times of the year, especially in colder climates. Biosolids should not be applied to frozen or snow covered grounds, while farm fields are sometimes not accessible during the growing season. Therefore, it is often necessary to provide a storage capacity in conjunction with land application programs. Even when the timing is right (for e.g., before crop planting in agricultural applications) weather can interfere with the application. Spring rain can make it impossible to get application equipment into farm fields, making it necessary to store biosolids until weather conditions improve.

Another disadvantage of land application is potential public opposition, which is encountered most often when the beneficial use site is close to residential areas. One of the primary reasons for public concern is odor. In the worst case situations, municipalities or counties may pass ordinances which ban or restrict the use of biosolids. However, many successful programs have gained public support through effective communication, an absolutely essential component in the beneficial use of biosolids. Despite many positive impacts to the environment, land application can have negative impacts on water, soil, and air if not practised correctly. Negative impacts to water result from the application of biosolids at rates that exceed the nutrient requirements of the vegetation. Excess nutrients in the biosolids (primarily nitrogen compounds) can leach from the soil and reach groundwater. Runoff from rainfall may also carry excess nutrients to surface water. However, because biosolids are slow released fertilizers, the potential for nitrogen compounds to leach from biosolids amended soil is less than that posed by the use of chemical fertilizers. In areas fertilized by either biosolids or chemicals, these potential impacts are mitigated by proper management practices, including the application of biosolids at agronomic rates (the rate nutrients are used by the vegetation.) Maintenance of buffer zones between application areas and surface water bodies and soil conservation practices will minimize impacts to surface water.

Negative impacts to soil can result from mismanagement of a biosolids land application. Federal regulations contain standards related to all metals of concern and application of biosolids which meets these standards should not result in the accumulation of metals to harmful levels. Stringent record keeping and reporting requirements on both the federal and state level are imposed to prevent mismanagement.

Odors from biosolids applications are the primary negative impact to the air. Most odors associated with land application are a greater nuisance than threat to human health or the environment. Odor controls focus on reducing the odor potential of the biosolids or incorporating them into the soil. Stabilization processes such as digestion can decrease the potential for odor generation. Biosolids that have been disinfected through the addition of lime may emit ammonia odors but they are generally localized and dissipate rapidly. Biosolids stabilization reduces odor and usually results in an operation that is less offensive than manure application.

Overall, a properly managed biosolids land application program is preferable to the use of conventional fertilizers for the following reasons (1):

a. Biosolids are a recycled product, use of which does not deplete nonrenewable resources such as phosphorous.
b. The nutrients in biosolids are not as soluble as those in chemical fertilizers and are therefore released more slowly.
c. Biosolids appliers are required to maintain setbacks from water resources and are often subject to more stringent soil conservation and erosion control practices, nutrient management, and record keeping and reporting requirements than farmers who use only chemical fertilizers or manures.
d. Biosolids are closely monitored.
e. The organic matter in biosolids improves soil properties for optimum plant growth, including tilth, friability, fertility, and water holding capacity. They also decrease the need for pesticide use.

A joint policy statement of the US Department of Agriculture, the US Food & Drug Administration, and the US EPA states, "...the use of high quality biosolids coupled with proper management procedures, should safeguard the consumer from contaminated crops and minimize any potential adverse effect on the environment" (18).

5. DESIGN CRITERIA

Design criteria for land application programs address issues related to application rates and suitable sites. Design criteria for physical facilities (such as stabilization) that are part of land application programs are discussed in other chapters. Biosolids, site, and vegetative characteristics are the most important design factors to consider.

Biosolids must meet regulatory requirements for stabilization and metals content. In addition, nutrient content and physical characteristics, such as percent solids, are used to determine the appropriate application rate for the crop that will be grown and the soil in which the crops will be grown.

Site suitability is determined based on such factors as soil characteristics, slope, depth to groundwater, and proximity to surface water. In addition, many states have established site requirements to further protect water quality. Some examples include:

a. Sufficient land to provide areas of nonapplication (buffers) around surface water bodies, wells, and wetlands.
b. Depth from the soil surface to groundwater equal to at least one meter.
c. Soil pH in the range of 5.5–7.5 to minimize metal leaching and maximize crop growing conditions.

Site suitability is also influenced by the character of the surrounding area. While odors and truck traffic may not be objectionable in an agricultural area, both will adversely impact residential developments and community centers close to fields where biosolids are applied. The type of vegetation to be grown is also a design consideration. Vegetation, like soil characteristics, will generally not exclude biosolids application since most vegetation will benefit from the practice. However, the type of vegetation will impact the choice of application equipment, the amount of biosolids to be applied, and the timing of applications. The amount of biosolids that may be applied to a site is a function of the amount of nutrients required by the vegetation and the amount of metals found in the biosolids. Table 2 summarizes the application frequency, timing, and rates for various types of sites (1,17).

Another factor to be considered in designing a land application program is the timing of applications. Long period of saturated or frozen ground limit opportunities for application. This is an important consideration in programs using agricultural lands; applications must be performed at times convenient to the farmer and must not interfere with the planting of crops. Most applications of biosolids to agricultural land occur in the early spring or late fall. As a result, storage or an alternate biosolids management option must be available to handle biosolids when application is not possible. Forest land and reclamation sites allow more leeway in the timing of applications. In some areas of the United States, application can proceed year round.

Application is the most beneficial on agricultural land in late fall or early spring before the crop is planted. Timing is less critical in forest applications when nutrients can be incorporated into the soil throughout the growing period. Winter application is less desirable in many locales. Rangelands and pasturelands also are more adaptable to applications during various seasons. Applications can be made as long as ground is not saturated or snow covered and whenever livestock can be grazed on alternate lands for at least 30 d after the application. The timing of single applications in land reclamation programs is less critical and may be dictated by factors such as regulatory compliance schedule.

6. PERFORMANCE

In 1995, approx 54% of wastewater treatment plants managed biosolids through land application, an increase of almost 20% from information reported in 1993 (5,19). The

Table 2
Typical Biosolids Application Scenarios

Type of site/vegetation	Schedule	Application frequency	Application rate
Agricultural land			
Corn	April, May, after harvest	Annually	5–10 dry t/acre
Small grains	March–June, August, fall	Up to 3 times/yr	2–5 dry t/acre
Soybeans	April–June, fall	Annually	5–20 dry t/acre
Hay	After each cutting	Up to 3 times/yr	2–5 dry t/acre
Forest land	Year round	Once every 2–5 yr	5–100 dry t/acre
Range land	Year round	Once every 1–2 yr	2–60 dry t/acre
Reclamation sites	Year round	Once	60–100 dry t/acre

Source: US EPA.

vast majority of these land application programs use agricultural land, with minor amount applied to forest land, rangeland, or land in need of reclamation.

The use of land application increased steadily in the 1980s for several reasons, including decreasing availability and increasing costs associated with landfill disposal. Research also helped refine procedures for proper land application. Meanwhile, implementation of the Nationwide Pretreatment Program (3) resulted in significant improvements in biosolids quality. The 1993 adoption of the Part 503 Rule created a structure for consistent application procedures across the nation. The regulations were developed with input from the US Department of Agriculture (USDA), the US Food and Drug Administration (USFDA), biosolids generators, environmental groups, the public, state regulators, and academic researchers. Conservative assumptions were used to create regulations to “protect public health and the environment from all reasonably anticipated adverse effects” (5).

Land application is a reliable biosolids management option as long as the system is designed to address such issues as storage or alternate management for biosolids during periods when application cannot take place due to unfavorable weather or field conditions. Public opposition rather than technical constraints is the most common reason for discontinuing land application programs (1). Martha Prothro, a Former Deputy Assistant Administrator for Water, US EPA stated that (20) *“In fact, in all the years that properly treated biosolids have been applied to the land, we have been unable to find one documented case of illness or disease that resulted.”*

Land application systems generally use uncomplicated, reliable equipment. Operations include pathogen reduction processing, dewatering, loading of transport vehicles, transfer to application equipment, and the actual application. Operations and maintenance considerations associated with pathogen reduction processing are discussed in refs. 12,16. The other operations require labor skills of heavy equipment operators, equipment maintenance personnel, and field technicians for sampling, all normally associated with wastewater treatment facilities. In addition, the biosolids generator is responsible for complying with state and local requirements as well as federal regulations. The biosolids manager must be able to calculate agronomic rates and comply with record keeping and recording requirements. In fact, the generator and land applier must sign certification statements verifying accuracy and compliance (1). The generator should also allocate time to communicate with farmers, landowners, and neighbors about the benefits of biosolids recycling. Control of odor, along with a viable monitoring program, is the most important for public acceptance (21). Detailed discussions and more information on biosolids recycling for land application can be found in refs. 22–27.

7. COST OF RECYCLING THROUGH LAND APPLICATION

It is difficult to estimate the cost of land application of biosolids without specific program details. For example, there is some economy of scale due to large equipment purchases. The same size machine might be needed for a program that manages 10 ton/d of dry biosolids as one managing 50 dry t/d; the cost of that machine can be spread over the 10 or 50 dry tons, greatly affecting average costs per dry ton. One source identified costs for land application varying from 60–290 USD/dry t in 1996 (28), which is equivalent to 71–344 in 2006 USD/t (Appendix; 29). This range reflects the wide variety in land application methods as well as varying methods to prepare biosolids for land application. For example, costs for programs using dewatered biosolids include an additional step, whereas, cost for programs using liquid biosolids do not reflect the cost of dewatering. They do, however, include generally higher transportation costs.

Despite the wide range of cost for land application programs, several elements must be considered in estimating the cost of any biosolids land application program (1):

a. Purchase of application equipment or contracting for application services.
b. Transportation.
c. Equipment maintenance and fuel.
d. Loading facilities.
e. Labor.
f. Capital, operation, and maintenance of stabilization facilities.
g. Ability to manage and control odor.
h. Dewatering (optional).
i. Storage or alternate management option for periods when application is not possible because of weather or climate.
j. Regulatory compliance, such as permit applications, site monitoring, and biosolids analyses.
k. Public education and outreach efforts.

Land must also be secured. Some municipalities have purchased farms for land application; others apply biosolids to privately held land. Some operating costs can be offset through the sale of the biosolids material. Since the biosolids reduce the need for fertilizers and pH adjustment, farmers pay to have biosolids applied to their lands.

8. BIOSOLIDS DISPOSAL ON LAND (LANDFILL)

Biosolids landfill can be defined as the planned burial of wastewater solids and processed biosolids at a designated land site. The solids are placed into a prepared site or excavated trench and covered with a layer of soil. The soil cover must be deeper than the depth of the plow zone (about 8–10 in.). For the most part, landfilling of screenings, grit, and ash is accomplished with methods similar to those used for biosolids landfilling (4).

9. BIOSOLIDS LANDFILL METHODS

Biosolids landfill methods can be grouped into three general categories:

a. Biosolids-only trench fill.
b. Biosolids-only area fill.
c. Co-disposal with refuse.

General site and design criteria are discussed under these categories. A detailed discussion of biosolids landfills can be found in the US EPA Technology Transfer Process Design Manual, Municipal Sludge Landfills (30), and in Office of Solid Waste Report, Disposal of Sewage Sludge into a Sanitary Landfill (31).

9.1. Biosolids-Only Trench Fill

Stabilized or unstabilized biosolids are placed within a subsurface excavation and covered with soil. Trench operations are more specifically categorized as follows: narrow trench and wide trench. Narrow trenches are defined as having width less than 10 ft; wide trenches are defined as having width more than 10 ft. The width of the trench is determined by the solids content of the receiving biosolids and its capability of supporting cover material and equipment. Distance between trenches should be large enough to provide sidewall stability, as well as space for soil stockpiles, operating equipment and haul vehicles.

Design considerations should include provisions to control leachate and gas migration, dust, vectors, and/or aesthetics. Leachate control measures include the maintenance of 2–5 ft of soil thickness between trench bottom and the highest groundwater level or bedrock (2 ft for clay to 5 ft for sand), or membrane liners and leachate collection and treatment system. Installation of gas control facilities may be necessary if inhabited structures are nearby. A comparison of the three types of trench fill (32) is shown in Table 3.

9.1.1. Narrow Trenches

Trenches are defined as narrow when their width is less than 10 ft (3 m). Biosolids are disposed in a single application and a single layer of cover soil is applied on top. Trenches are usually excavated by equipment based on solid ground adjacent to the trench, and equipment does not enter the excavation. Backhoes, excavators, and trenching machines are particularly useful. Excavated material is usually immediately applied as cover over an adjacent biosolids-filled trench. Biosolids are placed in trenches either directly from haul vehicles, through a chute extension, or by pumping. The main advantage of a 2–3 ft narrow trench is its ability to handle biosolids with relatively low solids content (15–20%). Instead of sinking to the bottom of the biosolids, the cover soil bridges over the trench and receives support from undisturbed soil along each side of the trench. A 3–10 ft width is more appropriate for biosolids with solids content of 20–28%, which is high enough to support cover soil.

The application rates range from 1200 to 5600 yd^3 of biosolids/acre (2270–10,580 m^3/ha). Excavated material can be either used immediately to cover an adjacent biosolids-filled trench or stockpiled alongside and used to cover the trench from which it was removed. The surface soil cover thickness is about 4 ft (1.3 m).

Table 3
Comparison of Design Criteria for Narrow and Wide Trench Landfill

Design criteria	Narrow trench (<10 ft)	Wide trench (>10 ft)
Sludge solids content	15–20% for 2–3 ft widths, 20–28% for 3–10 ft widths	20–28% for land-based equipment, more than 28% for sludge-based equipment
Ground slopes	Less than 20%	Less than 10%
Cover soil thickness	2–3 ft for 2–3 ft widths, 3–4 ft for 3–10 ft widths	3–4 ft for land-based equipment, 4–5 ft for sludge-based equipment
Sludge application rate	1200–5600 yd^3/acre	3200–14,500 yd^3/acre
Equipment	Backhoe with loader, excavator, trenching machine	Track loader, dragline, scraper, track dozer

Source: US EPA.

9.1.2. Wide Trenches

Trenches are defined as wide when they have width more than 10 ft (3 m). Trenches are usually excavated by equipment operating inside the trench. Track loaders, draglines, scrapers, and track dozers are suitable. Excavated material is stockpiled on solid ground adjacent to the trench for subsequent application as cover material. If biosolids are incapable of supporting equipment, cover is applied by equipment based on solid undisturbed ground adjacent to the trench. A front-end loader is suitable for trenches up to 10 ft wide; a dragline is suitable for trench width up to 50 ft. If biosolids can support equipment, a track dozer applies cover from within the trench.

Biosolids are placed in trenches by one of the two methods; from haul vehicles directly entering the trench and haul vehicles dumping from the top of the trench. Dikes can be used to confine biosolids to a specific area in a continuous trench. Disposal in wide trenches is suitable for biosolids with solids content of 20% or more. The application rates range from 3200 to 14,500 yd^3 of biosolids/acre (6050–27,400 m^3/ha).

The surface cover thickness depends on the solids concentration of the biosolids. The covered biosolids will only be capable of supporting equipment when the solids concentration of the biosolids exceeds 25–30% and the biosolids have been topped with 3–5 ft (1–2 m) of soil. The wide trench method has two distinct advantages; it is less land-intensive than the narrow trench method and groundwater protection can be provided by liners. The use of liners permits deeper excavations. The primary disadvantage of the wide trench method is the need for biosolids concentrations of more than 20% solids. Biosolids with solid contents of more than 30–35% will not flow, and extra effort is therefore required to spread them evenly in the trench. After maximum settlement has occurred in approximately one year, the area should be regraded to ensure proper drainage.

9.2. Biosolids-Only Area Fill

In the biosolids-only area fill method, the biosolids are mixed with soil and the mixture is placed on the original ground surface. This method requires substantial amount of imported soil but may be suitable in area where groundwater is shallow (liners can be easily installed) or bedrock prevails (i.e., where excavation is neither possible nor required). Stabilized biosolids are best-suited for this method, since daily cover is not usually provided.

To achieve stability and soil bearing capacity, sludge is mixed with a bulking agent, usually soil. The soil absorbs excess moisture from the sludge and increases its workability. The large quantities of soil required may require hauling from elsewhere. Provisions must be made to keep the stockpiled soil dry. Installation of a liner is generally required for groundwater control. Provisions are made for surface drainage control to prevent contamination of nearby surface waters, gas migration, dust, vectors and/or aesthetics. A comparison of the three types of area fill (32) is shown in Table 4.

9.2.1. Area Fill Mound

Area fill mound applications are generally suitable for stabilized biosolids with solids concentrations of 20% or more. Biosolids are mixed with a bulking agent, usually soil, and the mixture is hauled to the filling area, where it is stacked in mounds approx 6 ft high. Cover material is then applied in a 3 ft thickness. This cover thickness may be increased to 5 ft if additional mounds are applied on the top of the first lift. The appropriate sludge/soil bulking ratio and soil cover thickness depend upon the solids content of the sludge as received, the need for mound stability and bearing capacity as dictated by the number of lifts and equipment weight. Lightweight equipment with swamp pad tracks is appropriate for area fill mound operations; heavier wheel equipment is appropriate in transporting bulking material to and from stockpiles. A level area is required for disposal; however, the use of earthen containment structures permits disposal in hilly areas.

9.2.2. Area Fill Layer

Area fill layer applications are suitable for stabilized biosolids with solids as low as 15%. Soil is mixed with biosolids, either at the filling area or at a special mixing area. The biosolids/soil mixture is spread in even layers of approx 1 ft (0.3 m) thick, and 3–5 ft (1–1.5 m) of soil is added for final cover. Lightweight equipment with swamp pad tracks is appropriate for area fill layer operations; heavier wheel equipment is appropriate for hauling soil. Slopes should be relatively flat to prevent sludge from flowing downhill. However, if sludge solids content is high and/or sufficient bulking soil is used, the effect can be prevented and layering performed on mildly sloping terrain.

9.2.3. Dike Containment

Dike containment applications require biosolids with solids content of 20% or more. This method is suitable for either stabilized or unstabilized biosolids. If the disposal site is level, earthen dikes are used on all four sides of the containment area. If the site is at the toe of the hill, only a partial diking is required. Access is provided to the top of the dike so that haul vehicles can dump biosolids directly into the containment. Depending on the type of equipment used, the interim cover will vary from 1 to 3 ft (0.3–1 m) and the final cover from 3–5 ft (1–1.5 m).

Cover material is applied either by a dragline based on solid ground on top of the dikes or by track dozers directly on top of the sludge, depending upon sludge bearing capacity. Usually, operations are conducted without the addition of soil bulking agents, but occasionally soil bulking is added. Typical dimensions: 50–100 ft wide, 100–200 ft

Table 4
Comparison of Design Criteria for Area Fill Mound, Area Fill Layer, and Diked Containment Landfill

Design criteria	Area fill mound	Area fill layer	Diked containment
Sludge solids content	More than 20%	More than 15%	20–28% for land-based equipment; more than 28% for sludge-based equipment
Sludge characteristics	Stabilized	Stabilized	Stabilized or unstabilized
Ground slopes	No limitation if suitably prepared	Level ground preferred	Level ground or steep terrain if suitably prepared
Bulking required	Yes	Yes	Occasionally
Bulking ratio soil: sludge	0.5–2 soil: 1 sludge	0.25–1 soil: 1 sludge	0–0.5 soil:1 sludge
Sludge application rate	3000–14,000 yd^3/acre	2000–9000 yd^3/acre	4800–15,000 yd^3/acre
Equipment	Track loader, backhoe with loader, track dozer	Track dozer, grader, track loader	Dragline, track dozer, scraper

Source: US EPA.

long, 10–30 ft deep. Although diked containment is an efficient disposal method from the standpoint of land use, it may necessitate controls for leachate outbreaks.

9.3. Co-Disposal With Refuse

The term co-disposal is used when municipal biosolids are disposed of at a refuse landfill. There are distinct trade-offs in using co-disposal method rather than the biosolids-only method. Biosolids can be disposed of in this manner if they are mixed with refuse or with soil. Mixing techniques are discussed in detail in the US EPA Office of Solid Waste Report, Disposal of Wastewater Biosolids into a Sanitary Landfill (31).

9.3.1. Biosolids/Refuse Mixture

Stabilized or unstabilized biosolids with solids content of three percent or greater are mixed with the refuse. Normally biosolids content is approx 10% of the biosolids/refuse mixture. The biosolids are applied on top of the refuse at the working face of the landfill. The biosolids and refuse are thoroughly mixed before they are spread, compacted, and covered with soil. An interim cover of approx 1 ft (0.3 m) and a final cover of 2 ft (0.6 m) is used. Application rates range from 500 to 4200 yd^3 of biosolids/acre (950–7900 m^3/ha).

9.3.2. Biosolids/Soil Mixture

In this operation, biosolids are mixed with soil and the mixture is used as cover for a refuse landfill. This method requires stabilized biosolids with at least 20% solids content. It promotes vegetation growth over completed landfill area without the use of fertilizer. However, it may cause odor, as the biosolids are not completely buried. A final soil cover could be added if necessary to eliminate this problem.

Some wastewater treatment biosolids may not be suitable for landfilling by any of the methods previously described. For landfilling by biosolids, the solids concentration should be 15% or more. Although soil may be used as a bulking agent to effectively increase the solids concentration to this level, cost-effectiveness might become a problem. Solids concentrations down to three percent are tolerated for co-disposal, but the absorptive capacity of the refuse should not be exceeded. An assessment of the suitability of various biosolids types is given in Table 5. In general, only stabilized and dewatered biosolids are recommended for landfill disposal.

9.4. Landfilling of Screenings, Grit, and Ash

Screenings and grit normally contain some putrescible materials and should be covered every day. Odor from temporarily uncovered solids may be alleviated by sprinkling the solids with lime. Special care should be exercised to assure vector control (e.g., safe poisons for rodent control, spraying for flies, and animal-proof fencing to keep pets from the area). Residues (ash from the combustion of municipal wastewater solids) generally contain high concentrations of trace metals. Leachate from sites where incinerator ash is landfilled must be controlled to prevent metals contamination of groundwater.

10. PRELIMINARY PLANNING

The purpose of the preliminary planning activity is to select a disposal site and suitable method(s) of disposal. Preliminary planning is followed by detailed design, initial site development, site operation and maintenance, and final site closure. Site selection is the major activity during the preliminary planning phase. As the selection of a site is not completely independent of the selection of a method, the preliminary planning phase should also include the determination of biosolids characteristics and the identification of alternate landfill methods for each site.

10.1. Biosolids Characterization

Biosolids must be characterized as to quantity and quality. An estimate of the average biosolids quantity is necessary to establish landfill area requirements and the probable life of the disposal site. Data on minimum and maximum biosolids quantities is important for developing an understanding of daily operating requirements. Maximum daily biosolids quantities will govern equipment and storage facility sizing and daily operating schedule. The character of the biosolids to be landfilled is directly related to the choice of a landfill method. Biosolids quality and the corresponding leachate can be roughly correlated; design of leachate treatment facilities is more effective if biosolids quality is known.

Parameters that should be analyzed are discussed briefly below (a-h) (4). Although all of these may not be critical to the design of a particular disposal system, a complete analysis is necessary, because the biosolids must be adequately characterized.

a. **Concentration.** Concentration or solids content of biosolids is related to the nature of wastewater treatment and biosolids processing steps. The type and operation of dewatering equipment may have a significant effect on the biosolids concentration. A certain degree of flexibility should be incorporated into the design of landfills to compensate for the variability in solids concentration of dewatered biosolids.

Table 5
Suitability of Biosolids for Landfill

	Sludge only landfilling		Codisposal landfilling	
Type of sludge	Suitability	Reason	Suitability	Reason
Liquid–unstabilized				
Gravity thickened primary, WAS and primary, and WAS	NS	OD, OP	NS	OD, OP
Flotation thickened primary and WAS, and WAS without chemicals	NS	OD, OP	NS	OD, OP
Flotation thickened WAS with chemicals	NS	OP	NS	OD, OP
Thermal conditioned primary or WAS	NS	OD, OP	MS	OD, OP
Liquid–stabilized				
Thickened anaerobic digested primary and primary, and WAS	NS	OP	MS	OP
Thickened aerobic digested primary and primary, and WAS	NS	OP	MS	OP
Thickened lime stabilized primary and primary, and WAS	NS	OP	MS	OP
Dewatered–unstabilized				
Vacuum filtered, lime conditioned primary	S	–	S	–
Dewatered–stabilized				
Drying bed digested and lime stabilized	S	–	S	–
Vacuum filtered, lime conditioned digested	S	–	S	–
Pressure filtered, lime conditioned digested	S	–	S	–
Centrifuged, digested and lime conditioned digested	S	–	S	–
Heat dried				
Heat dried digested	S	–	S	–
High temperature processed				
Incinerated dewatered primary and primary, and WAS	S	–	S	–
Wet-air oxidized primary and primary, and WAS	NS	OD, OP	MS	OD, OP

WAS, waste-activated sludge; NS, not suitable; MS, marginally suitable; S, suitable; OD, odor problems; OP, operational problems.

Source: US EPA.

b. **Volatile content.** Volatile solids are a measure of the organic content present in the solid fraction of biosolids. This organic matter is eventually broken down into methane gas and other digestion by-products. Typically, volatile solids represent 60–80% of the total solids in raw primary biosolids and 30–60% in anaerobically digested primary solids.

c. **Nitrogen.** Nitrogen found in biosolids is a potential source of groundwater pollution. The total quantity and type of nitrogen are of importance. Nitrate is relatively mobile in soil and is therefore of concern.
d. **Inorganic ions.** Inorganic ions such as heavy metals are found in most municipal biosolids. These are more readily leached if soil and biosolids are acidic. If near neutral or alkaline conditions are maintained, the metals will not be as readily leached from the biosolids or through the soil.
e. **Bacteriological quality.** Biosolids treatment systems reduce the number of pathogens (8,10–12) and the possibility of pathogenic contamination associated with landfilling of biosolids.
f. **Toxic organic compounds.** Toxic organic compounds can present potential contamination problems. Solids contaminated with toxic materials must be placed in appropriately designated disposal facilities.
g. **pH.** Acidic conditions promote leaching of heavy metals and other compounds from the biosolids.

10.2. Selection of a Landfilling Method

Relationship among the characteristics of alternative landfill sites, the characteristics of the biosolids to be landfilled, and the landfill method needs to be considered in the preliminary planning process. These relationships are summarized in Table 6.

10.3. Site Selection

Site selection is a critical process in the planning of a biosolids landfill project. It is directly related to the method of ultimate disposal. The site finally selected must be suitable for the type of biosolids to be disposed off and situated in a convenient, yet unobtrusive, location.

10.3.1. Site Considerations

The following factors must be considered during the evaluation of possible landfill sites. Information on these factors should therefore be collected and assessed in advance of the final decision making process.

a. **Haul distance.** The most favorable haul conditions combine level terrain and minimum distances.
b. **Site life and size.** The site life and size are directly related to the quantity and characteristics of the biosolids and the method used for landfilling. Since the entire site cannot be used as fill area, both the gross area and the usable or fill area must be considered in determining the site size. Initially, the life of the site can be estimated. As the landfill is used, the expected life should be re-evaluated to ensure adequate capacity for future operations.
c. **Topography.** In general, biosolids landfilling is limited to sites with minimum slopes of one percent and maximum slopes of 20%. Flat terrain tends to result in ponding, whereas steep slopes erode.
d. **Surface water.** The location and extent of surface water in the vicinity of the landfill site can be a significant factor in the selection process. Existing surface water and drainage near proposed sites should be mapped and their present and proposed uses outlined. Leachate control measures including collection and treatment may be required as part of the landfill design.
e. **Soils and geology.** Soil is an important determinant in the choice of an appropriate biosolids landfilling site. Properties such as texture, structure, permeability, pH, and cation exchange capacity, as well as the characteristics of soil formation, might influence the selection of the site. The geology of possible landfill sites should be thoroughly examined

Table 6
Biosolids and Site Conditions

Method	Sludge solids content (%)	Appropriate sludge characteristics	Appropriate hydrogeology	Appropriate ground slope
Narrow trench	15–28	Unstabilized or stabilized	Deep groundwater and bedrock	<20%
Wide trench	≥20	Unstabilized or stabilized	Deep groundwater and bedrock	<10%
Area fill mound	≥20	Stabilized	Shallow ground water or bedrock	Suitable for steep terrain as long as level area is prepared for mounding
Area fill layer	≥15	Unstabilized or stabilized	Shallow ground water or bedrock	Suitable for medium slopes but level ground preferred
Diked containment	≥20	Stabilized	Shallow groundwater or bedrock	Suitable for steep terrain as long as a level area is prepared inside dikes
Sludge/refuse mixture	≥3	Unstabilized or stabilized	Deep or shallow groundwater or bedrock	<30%
Sludge/soil mixture	≥20	Stabilized	Deep or shallow groundwater or bedrock	<5%

Source: US EPA.

to identify any fault, major fractures and joint sets. The possibility of aquifer contamination through irregular formations must be studied.

f. **Groundwater.** Data on groundwater in the vicinity of potential landfill sites is essential to the selection process. Knowledge of characteristics such as the depth to groundwater, the hydraulic gradient, the quality and use of the groundwater, and the location of recharge zones is essential for determining the suitability of a potential landfill site.

g. **Vegetation.** The type and quantity of vegetation in the area of proposed landfill sites should be considered in the evaluation. Vegetation can serve as a natural buffer, reducing visual impact, odor, and other nuisances. At the same time, clearing a site of timber or other heavy vegetation can add significantly to the initial project costs.

h. **Meteorology.** Prevailing wind direction, speed, temperature and atmospheric stability should be evaluated to determine potential odor and dust impacts downwind of the site.

i. **Environmentally sensitive areas.** Environmentally sensitive areas such as wetlands, flood plains, permafrost areas, critical habitats of endangered species, and recharge zones of aquifers should be avoided when selecting a landfill site (5,6).

j. **Archaeological and historical significance.** The archaeological and historical significance of proposed sites should be determined early in the evaluation process. Any significant finds at the selected site must be accommodated prior to final approval.

k. **Site access.** Haul routes should be major highways, or arterials, preferably those with a minimum traffic during normal transport hours. Proposed routes should be studied to determine impacts on local use and the potential effects of accidents. Transport through nonresidential area is preferable to transport through residential area, high-density urban areas, and areas with congested traffic. The access roads to the site must be adequate for the anticipated traffic loads.

l. **Land use.** Zoning restrictions and future development on potential sites should be considered in the selection process. Ideally, the biosolids landfill site should be located on land considered unsuitable for higher uses; however, the designer should be aware that this may be a politically sensitive issue and maximum public participation must be assured.

m. **Costs.** Cost-effectiveness of each potential landfill site must be evaluated. Factors to be included in the economic evaluation include capital costs and operating and maintenance (O&M) costs. In the latter category, biosolids hauling may prove to be a significant component. The trade-offs between high capital and high O&M costs will depend on the design life of the landfill. These trade-offs will become evident when the total annual (amortized capital and O&M costs) are compared.

10.3.2. Site Selection Methodology

The selection procedure can be roughly divided into three phases:

a. Initial inventory and assessment of sites.
b. Screening of potential sites.
c. Final site selection.

Initial inventory and assessment is designed to develop a list of potential sites that can be evaluated and rapidly screened to produce a manageable number of candidate sites. Information used in this phase is generally available and readily accessible. Investigation of each option becomes more detailed as the selection procedure progresses. Initial assessments will consist of identifying federal, state, and local regulatory constraints, eliminating inaccessible areas, locating potential sites, roughly assessing the economic feasibility of such sites, and performing preliminary site evaluations. The less desirable sites are eliminated on the basis of preliminary economics, regulatory, and technical information. A public participation program is initiated (33). Attitudes of the public should be determined early. The public may assist in identifying candidate sites.

Sites remaining after the initial assessment are subjected to closer scrutiny. Information used in evaluating each option is more detailed and somewhat more site-specific than in the initial assessment. Remaining sites may be rated by a scoring system including both objective and subjective evaluations. Candidate systems with the lowest overall ratings are eliminated, and the higher rated systems are carried forward for final evaluation.

Site selection findings for the remaining candidate systems should provide input into an environmental impact report, if required. Public attitudes toward the remaining sites should also be determined. Methodology for final site selection is similar to that for the screening procedure just discussed, in which rating systems are still used. However, each site remaining is investigated in greater detail. Public hearings may also be scheduled so that final inputs can be received from local government officials and the public.

Once the best sites are determined, they must be acquired. Site acquisition should begin immediately following acceptance of the program by local, state, and federal regulatory authorities. The several acquisition procedures include: purchase option, outright purchase, lease, condemnation and/or other court action, and land dedication. It will generally prove advantageous to purchase the site rather than hold a long-term lease. The managing agency's responsibility will normally extend well beyond the life of the site. Certain advantages may also be gained by leasing with an option to buy the site at the time of planning approval. A purchase option assures the availability of land upon completion of the facility planning process. This approach also allows time for the previous owner to gradually phase out operations, if desired.

11. FACILITY DESIGN

11.1. Regulations and Standards

Local, State, and Federal regulations and standards must be fully understood before the landfill is designed. Consideration must be given to requirements governing the degree of biosolids stabilization, the loading rates, the frequency and depth of cover, monitoring, and reporting (5,6). The design should conform to all building codes and should include adequate buffer zones to protect public roads, private structures, and surface water.

US EPA Rule Part 503 (5,6) states that in case landfill sites use liners (hydraulic conductivity $\geq 1 \times 10^{-7}$ cm/s) and leachate collection system there are no pollutant concentration limits because pollutant leaching will be collected and treated. Where landfill sites are with no liners, limits on three pollutants (arsenic, chromium, and nickel) are established. While these vary based on the distance of the active landfill boundary from the site property line, the most extreme values allowed are listed in Table 7.

The Rule also requires that the landfill operation does not cause the maximum contaminant level for nitrates in groundwater to be exceeded or to cause the existing level of nitrates to be exceeded if it already exceeds the maximum contaminant level. Part 503 also requires that the biosolids be either of class A or class B with respect to pathogen control unless the biosolids are covered daily with soil or other material. It must be stated that in many locations state regulations may be more strict even requiring a liner system.

Obtaining permits for construction and operation of biosolids landfills can be a long and costly process. To minimize delays associated with this task, permit application should be initiated early in the design stage. A sound regulatory-consultant relationship and a mutual understanding should be developed.

The following is a partial list of the permits which may be required (4):

a. US EPA special permit if landfill is in wetlands or other sensitive areas (5,6).
b. Army Corps of Engineers permit for construction of levees, dikes, or containment structures to be placed in the water in a wetlands area.
c. Office of Endangered Species permit if landfill is located in critical habitat of an endangered species.
d. Solid waste management permit.
e. Special use permit.
f. Highway department permit.
g. Construction permit.
h. Building permit.
i. Drainage and/or flood plain alteration permit.

11.2. Site Characteristics

Site characteristics should be clearly described and analyzed to ensure the suitability of the landfill site and the method of landfilling. Design phase work will build upon

Table 7
Maximum Allowable Pollutant Concentrations^a in Biosolids for Disposal in Landfills—No Liner and Leachate Collection systems

Distance from active landfill boundary to property line (m)	Arsenic (mg/kg)	Chromium (mg/kg)	Nickel (mg/kg)
0 to <5	30	200	210
25 to <50	34	220	240
50 to <75	39	260	270
75 to <100	46	300	320
100 to <125	53	360	390
125 to <150	62	450	420
≥150	73	600	420

Source: US EPA.
[a] All pollutant concentrations are dry-weight basis.

planning phase data but will be carried to a higher level of detail and include working drawings.

1. **Site plan.** The site plan should contain the following minimum information:
   a. Boundaries of fill area and buffer zones.
   b. Topographic features and slopes of fill area and buffer zones.
   c. Location of surface water, roads, and utilities.
   d. Existing and proposed structures and access roads.
   e. Vegetation to remain and to be removed; areas to be vegetated.
2. **Soils.** The soil characteristics at the landfill site should be thoroughly catalogued and mapped. The information of most importance to the design and operation of the landfill includes depth, texture, structure, bulk density, porosity, permeability, moisture, stability, and ease of excavation. Areas with rocky soil or extensive rock outcrops should be noted. The pH and cation exchange capacity have a direct bearing on heavy metal transport through the soil. Translocation of metals must be considered to ensure protection of surface and groundwater supplies.
   a. *Groundwater.* The groundwater aquifers underlying the landfill site must be located. Depth of the aquifer under varying conditions should be determined at several locations. Other characteristics such as the direction and rate of flow, the hydraulic gradient, the quality, and present and planned uses should also be established. Location of the primary recharge zones is critical in protecting quality.
   b. *Subsurface geology.* The geological formations underlying the landfill are important in establishing the design parameters. Critical design parameters include the depth, distribution, and characteristics of subsurface soils in relation to stability and groundwater transmissibility.
   c. *Climate.* Climate can influence many factors in the design of landfills. Climatic conditions affect the rate of organic decomposition, the composition and quantity of leachate and runoff, the day-to-day fill operations, and the dispersion of odors and dust. Information such as seasonal temperature, precipitation, evaporation, wind direction and speed and atmospheric stability, can be obtained from a local weather station.
   d. *Land use.* The present and proposed use of the landfill site and adjacent properties should be evaluated. If the site is already dedicated to refuse or biosolids disposal, it is unlikely that expanding it will result in adverse impacts. However, if the site is located in or near

a populated area, extensive control measures may be needed to eliminate concerns and minimize any public nuisance which would detract from the value of adjacent properties.

11.3. Landfill Type and Design

More than one biosolids landfill method may be suitable for the selected site, as shown in Table 2. If this is the case, a method must be selected before the final design is begun. Maximizing utilization of the site is an important consideration in method selection. If daily cover is to be applied, the daily biosolids generation rate will affect the net capacity of the site. If several days are required to fill a trench, as the result of low biosolids generation, and cover is required each day, then the ratio of biosolids/cover will be less than for sites managing larger biosolids quantities. The net biosolids capacity will be higher at sites where trenches are filled each day.

The amount by which the net capacity of the site will be reduced will vary with the landfill methods, the specific site, and the daily biosolids generation rate. Before a final method is selected, estimates of net capacity and site life should be made for each. Additional design criteria are summarized in Table 8.

11.4. Ancillary Facilities

Ancillary facilities may be needed in association with the landfill site. These are described briefly in the following sections.

a. **Leachate controls.** Leachate from the landfill site must be contained and treated to eliminate potential water pollution and/or potential public health problems. In many cases, leachate containment and treatment may be required by state or local regulations. Numerous methods are available for controlling leachate, including drainage, natural attenuation, soil or membrane liners, or collection and treatment (34–42). The method and the design features chosen are specific for each project. Table 9 depicts biosolids-only leachate quality for one site sampled over 2 yr.
b. **Gas control.** Gas produced by decomposition of organic matter is potentially dangerous. This condition is of particular concern if the landfill is located near a populated area. Methane gas, in particular, is highly explosive if confined in an enclosed area. Control of the gases produced at the landfill must be provided. Two widely accepted methods control paths of gas migration. Permeable methods usually consist of a gravel-filled trench around the fill area for intercepting migrating gas and venting it to the atmosphere. Impermeable methods consist of placing a barrier of low permeability material, such as compacted clay, around the fill area to minimize lateral movement of gas. This method provides for gas venting through the cover material. In general, methane recovery is not cost-effective at biosolids-only or small co-disposal sites.
c. **Roads.** Paved access and on-site roads are necessary at the landfill site. Temporary roads may be constructed of well compacted natural soil or gravel. Considerations should include grades, road surface and stability, and climate. Grades in excess of ten percent should be avoided. Provisions should be made to allow trucks to turn around within the site area.
d. **Soil stockpiles.** Storage area should be provided for on-site stockpiling of transported soil where on-site soil is insufficient or their use inappropriate. The quantity and type of soil to be stockpiled depends on the individual demands of the landfill. Stockpiles may also be desirable for winter operations where frozen ground may limit excavation.
e. **Inclement weather areas.** Special landfill areas should be placed near the entrance to the site so that operations may be continued during inclement weather. Paved or all-weather roads should be provided for working these sites.

Table 8
Landfill Design Criteria

Method	Sludge solids content %	Trench width (ft)	Bulking required	Bulking agent	Bulking ratio^a	Cover thickness (ft)		Imported soil required	Sludge application rate (yd^3/acre^b)	Equipment
						Interim	final			
Sludge only trench fill										
Narrow trench	15–20^c	2–3	No	–	–	–	2–3	No	1200–5600	Backhoe with loader, excavator, trenching machine
	20–28^c	3–10	No	–		–	3–4			
Wide trench	20–28^c	10	No	–	–	–	3–4	No	3200–14,500	Track loader, dragline, scraper, track dozer
	≥28^d	10	No	–		–	4–5			
Sludge only-area fill										
Area fill mound	≥20^{c,d}	–	Yes	Soil	0.5–2 soil: 1 sludge	3	3–5	Yes	3000–14,000	Track loader, backhoe with loader, track dozer
Area fill layer	≥15^d	–	Yes	Soil	0.25–1 soil: 1 sludge	0.5–1	2–4	Yes	2000–9000	Track dozer, grader, track loader
Diked containment	20–28^c	–	No	Soil	0.25 soil: 1 sludge		3–4	Yes	4800–15,000	Dragline, track dozer, scraper
	≥28^d		No	Soil						

(Continued)

Table 8 (*Continued*)

Method	Sludge solids content %	Trench width (ft)	Bulking required	Bulking agent	Bulking ratio[a]	Cover thickness (ft) Interim	Cover thickness (ft) final	Imported soil required	Sludge application rate ($yd^3/acre^b$)	Equipment
Codisposal with refuse										
Sludge/ refuse mixture	$\geq 3^d$	–	Yes	Refuse	4–7 t refuse: 1 wet ton sludge	0.5–1	2	No	500–4200	Dragline, track dozer
Sludge/soil mixture	$\geq 20^d$	–	Yes	Soil	1 soil: 1 sludge	0.5–1	2	No	1600	Tractor with disc, grader, track loader

Source: US EPA.

[a] Volume basis unless otherwise noted.

[b] In actual fill areas.

[c] Land-based equipment.

[d] Sludge-based equipment.

[e] But sometimes used.

1 ft, 0.305 m; 1 yd^3, 0.765 m^3; 1 acre, 0.405 ha.

Table 9
Leachate Quality From Biosolids[a]-Only Landfill

Constituents	Values[b]
pH	6.7
TOC	1000[c]
COD	5100[d]
Ammonia nitrogen	198[d]
Nitrate nitrogen	0.28[d]
Chloride	6.7
Sulfate	10
Specific conductivity	3600[e]
Cadmium	0.017
Chromium	1.1
Copper	1.3
Iron	170
Mercury	0.0004
Nickel	0.31
Lead	0.60
Zinc	5

Source: US EPA.

[a] Data from "Site 8" monitored from July 1975 through September 1977. First received sludge in 1973. Receives unstabilized primary and WAS, gravity thickened and centrifuged. Sludge is lagooned, allowed to dry, and covered with soil. Soil characteristics: sand and gravel, glacial deposites.

[b] Specific conductivity in micromhos/cm, pH in units, all others in mg/L.

[c] Ranged from 1 to 3000 mg/L.

[d] Limited to early part of sampling program.

[e] Ranged from 340 micromhos/cm to 10,000 micromhos/cm.

f. **Structures.** An office and employee facilities should be located at the landfill site. For large operations, a permanent structure should be provided. At smaller sites a trailer might suffice. An equipment barn and shop may be desirable for some locations.
g. **Utilities.** Electrical, water, communication and sanitary services should be provided for large landfill operations. Chemical toilets, bottled water, and on-site electrical generation may reduce the cost of obtaining services from utility companies. This approach may be appropriate for remote sites.
h. **Fencing.** The landfill site should be fenced. Access should be limited to one or two secured entrances. The height and type of fence should suit local conditions. A 6-ft (1.8 m) chain link fence topped with barbed wire will restrict trespassers; a wooden fence or hedge is effective for screening the operation from view, and a 4-ft (1.2 m) barbed wire fence will keep cattle or sheep away from the site area.
i. **Lighting.** Portable lighting should be provided if landfill operations are carried out at night. Permanent lights should be installed for all structures and heavily used access roads.
j. **Wash racks.** A cleaning program should be required for frequently used equipment. A curbed wash pad and collection basin should be provided to contain the contaminated washwater for treatment.

k. **Monitoring wells.** It is crucial to monitor groundwater. The number, type, and location of monitoring wells and monitoring frequency should be designated to meet specific conditions associated with the landfill.

l. **Landscaping.** Depending on the size and location of the landfill, landscaping may be an important design factor. The aesthetic acceptability of the landfill is critical, especially in an urban or densely populated area. In general, shrubbery chosen should require little maintenance and become an effective visual barrier.

11.5. Landfill Equipment

A wide variety of equipment may be required for a biosolids landfill. The type of equipment depends on the landfill method employed and on the quantity of biosolids to be disposed of. Equipment will be required for biosolids handling, excavation, backfilling, grading, and road construction. Table 10 presents typical equipment performance characteristics for various biosolids landfilling methods.

11.6. Flexibility, Performance, and Environmental Impacts

Because biosolids characteristics and quantities may change, a landfill site should be designed with maximum flexibility. As the life of a landfill is difficult to accurately predict, expansion may be needed sooner than originally planned or it may be delayed. Any change in wastewater treatment or biosolids management processes may affect the nature and quantity of biosolids produced. Operational modifications may be needed if these changes are drastic. The landfill design should be such that changes can be made without major disruption to operations.

Reliability is another important factor in designing a landfill operation. Operation should continue even in inclement weather. Special work areas and storage facilities should be available on site for emergency operations or unexpected equipment failures. Although the overall performance of a biosolids landfill may be difficult to predict accurately, certain operating parameters should be estimated. The site life depends on many factors; an estimate is needed for purposes of economic evaluations and future planning. Biosolids application rate and soil cover requirements should be estimated before scheduling initial operations. Performance can be more closely predicted after actual operating experience is gained. Specific areas of environmental impact vary among landfill locations. Crucial impact areas include: traffic, land use, air quality, surface and groundwater quality, public health, aesthetics, wildlife, and habitats of endangered species. Adverse impacts should be mitigated during the site selection process or by specific measures in the design.

12. OPERATION AND MAINTENANCE

A biosolids landfill should be viewed as an ongoing construction site. However, unlike conventional construction, the operating parameters of a biosolids landfill often change and may require innovative alterations and contingency plans. An effective landfill requires a detailed operational plan. Equipment selection should be compatible with biosolids characteristics, site conditions, and landfill method. Operational procedures can be separated into those specific to the landfill method and those applicable to biosolids landfills in general. Method-specific procedures include: site preparation, biosolids unloading, biosolids management and covering. General procedures include

Table 10
Landfill Equipment Performance Characteristics

			Equipment type									
Landfill method	Submethod	Equipment function	Trenching machine	Backhoe with loader	Excavator	Track loader	Wheel loader	Track dozer[a]	Scraper	Dragline	Grader	Tractor with disc
Trench	Narrow trench	Trench construction	G	G	G	–	–	–	–	G	–	–
		Covering	G	G	F	G	G	G	–	G	–	–
	Wide trench	Trench construction	–	–	–	G	F	G	G	G	–	–
		Covering	–	–	–	F	–	G	–	G	–	–
Area fill	Mound	Soil hauling	–	F	–	F	G	–	G	–	–	–
		Mixing	–	F	–	G	F	G	–	–	–	–
		Sludge hauling	–	F	–	F	G	–	F	–	–	–
		Mounding	–	G	–	G	F	F	–	–	–	–
		Covering	–	F	–	G	F	G	–	G	–	–
	Layer	Soil hauling	–	F	–	F	G	–	G	–	–	–
		Mixing	–	F	–	G	G	G	–	–	–	–
		Sludge hauling	–	F	–	F	G	–	F	–	–	–
		Layering	–	–	–	F	–	G	G	–	G	–
		Covering	–	–	–	F	–	G	G	–	G	–
	Diked containment	Soil hauling	–	F	–	F	G	–	G	–	–	–
		Dike construction	–	–	–	F	F	G	G	–	–	–
		Covering	–	–	–	–	–	G	–	G	–	–

(Continued)

Table 10 ***(Continued)***

			Equipment type									
Landfill method	Submethod	Equipment function	Trenching machine	Backhoe with loader	Excavator	Track loader	Wheel loader	Track dozer[a]	Scraper	Dragline	Grader	Tractor with disc
Codisposal	Sludge/refuse	Spreading	–	–	–	F	–	G	–	–	–	–
		Covering	–	–	–	F	–	G	F	–	–	–
	Sludge/soil	Sludge spreading	–	–	–	F	–	G	–	–	F	–
		Mixing	–	–	–	–	–	F	–	–	–	G
		Hauling	–	–	–	G	F	–	F	–	–	–
		Covering	–	–	–	F	F	G	F	–	–	–

Source: US EPA.

G, good (fully capable of performing function listed. Equipment could be selected solely on the basis of function listed); F, fair (marginally capable of performing function listed. Equipment should be selected on the basis of full capabilities in other function); –, not applicable (cannot be used for function listed).

[a] Caterpillar D-6 generally is the largest track dozer appropriate for a sludge landfill although some engineers are investigating the use of the Caterpillar LG-T, double-wide track dozer.

scheduling, equipment selection and maintenance, management and reporting, safety, and environmental control. These procedures are discussed in detail in Municipal Biosolids Landfill (30) and in Sanitary Landfill Design and Operation (31). Important points are summarized next.

12.1. Operations Plan

As with any construction activity, biosolids landfilling must proceed according to detailed plans and operating schedules. The operation plan should address all relevant method-specific or general operating procedures for the landfill, including:

a. Hours of operation.
b. Measuring procedures.
c. Traffic flow and unloading procedures.
d. Special wastes handling.
e. Inclement weather operations.
f. Environmental monitoring and control practices.

An operation plan is an important tool for providing continuity of activities, monitoring and control of progress, and personnel training.

12.2. Operating Schedule

Major features of the operating schedule include:

a. Hours of operation,
b. Availability of qualified personnel,
c. Site preparation schedule, and
d. Equipment maintenance schedule.

The hours of operation must be such that the site is open when biosolids are to be received. If variations in the rate of receipt are expected during the day, it may be desirable to schedule for equipment and personnel accordingly. The schedule may need to provide for the application of daily soil cover.

12.3. Equipment Selection and Maintenance

Equipment selection depends largely upon the landfill method, design dimensions, and biosolids quantity. Selection must be based upon the functions to be performed and the cost of alternate machines. Table 10 summarized general selection criteria. Table 11 presents examples of equipment choices for seven landfill schemes.

Equipment maintenance can be more expensive than the amortized annual purchase cost. A scheduled preventive maintenance program should be followed to control maintenance costs. Operators should perform routine daily maintenance (e.g., check fluid levels, cleaning, and so on). The operating schedule should provide period for thorough maintenance.

12.4. Management and Reporting

Management and reporting activities include the maintenance of activity records, performance records, required regulatory reports, cost records, on-site supervision and public relations activities (43–48). Activity records include equipment and personnel accounts, biosolids and (if applicable) solid waste receipts, cover material quantities,

Table 11
Typical Equipment Type and Number as a Function of Landfill Method and Site Loading

	Trench method										Area fill method															Codisposal method[a]									
	Narrow trench					Wide trench					Mound					Layer					Diked containment					Sludge/ refuse					Sludge/ soil				
Equipment	1^b	2^c	3^d	4^e	5^f	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5	1	2	3	4	5
Trenching machine				1	2																														
Backhoe with loader	1	1		1^g	1							1^g	1^g	1^g	1																				
Excavator			1																																
Track loader						1	1^g	1		1^g	1	1	1	1	1															1			1^g	1^g	2^g
Wheel loader														1	1					1^g															
Track dozer		1^g	1	1	2^g		1^g	1	1	2^g			1^g	1	1	1	1	1	2^g	2	1	1^g	1^g	1	2^g			1^g	1	1					
Scraper									1^g	1			1^g	1^g	1		1^g	1^g	1^g	1			1^g	1	1										
Dragline																						1	1	1	1										
Grader																																		1^g	1
Tractor with disc																															1	1	1^g	2^g	2
Total	1	2	2	3	5	1	2	2	2	4	1	2	4	5	5	1	2	2	3	4	1	2	3	3	4	–	–	1	1	2	1	1	2	4	5

Source: US EPA.

[a] Additional equipment only.

[b] Scheme 1–10 wet t/d.

[c] Scheme 2–50 wet t/d.

[d] Scheme 3–100 wet t/d.

[e] Scheme 4–250 wet t/d.

[f] Scheme 5–500 wet t/d.

[g] May not receive 100% utilization.

and used site area layouts. These records become bases for scheduling site development, gauging efficiency, and any billing as required.

Performance records may be required as a part of the regulatory process. Regulatory agencies may perform periodic inspections on a scheduled or an unscheduled basis. Operating and supervisory personnel must be aware of these requirements. For the purposes of safety and control, the site should be staffed with two or more persons. At smaller sites, where only one operator is required, daily visits or phone checks should be made.

12.5. Safety

Providing a safe working environment at the landfill site should be a part of general O&M, and certain safety features should be built into the design. Certain practices must be followed daily to provide safe working conditions. The operations plan should have a separate safety section, as well as specific safety guidelines for each operation and feature of the landfill.

a. **Soil and fill stability.** The stability of the soil and fill can present a critical safety problem, particularly with the use of large equipment. Disturbed and filled areas should be approached cautiously as should muddy areas or areas subject to erosion.
b. **Equipment operation.** The operation of large, earth-moving equipment presents the potential for accidents. Only fully trained operators should be allowed to use such equipment. Regular maintenance and safety checks can greatly reduce the number of accidents associated with equipment failure and operator error.
c. **Gas control.** Caution must be used when dealing with gas control equipment. The O&M manual should contain a complete set of instructions on the safe servicing of gas control and monitoring equipment, and the operation of this equipment should be explained periodically at operation and safety training sessions.

12.6. Environmental Control

The protection of the environment and public health are important aspects of the landfill operation. The operation plan should contain guidelines for providing this protection and actual operations should conform to the guidelines. Potential environmental problems and the requirements for their control are summarized in Table 12. Critical areas are discussed next.

a. **Environment.** Environmental protection is generally focused on leachate and runoff controls for preventing surface and groundwater contamination. Trench liners must be kept intact during and after tilling operations (49). Drainage systems should be checked to see that they are functioning as designed. If monitoring indicates that adverse environmental impacts are occurring or pending, immediate corrective action should be taken.
b. **Public health.** Protection of public health should be a foremost concern in the operation of biosolids landfills. Protection of water supplies and particularly water aquifers is an obvious responsibility. In addition, control of potential disease by reduction of vectors, the adequate venting of explosive or toxic gases and the restriction of access to the landfill site are the responsibility of the operators.
c. **Social welfare.** Minimizing the negative aesthetic impacts of a biosolids landfill can greatly increase public acceptance. Control of odor, noise, and other nuisances are generally straight-forward and should be accomplished as part of the daily operating routine. Efforts should be made to reduce the undesirable social impacts of the fill operation.

Table 12
Potential Environmental Problems and Control Practices

	Environmental problems									
Control practice	Spillage	Siltation and erosion	Mud	Dust	Vectors	Odors	Noise	Aesthetics	Health	Safety
Safety program										X
Maintain washrooms for personnel									X	
Training of new personnel	X	X	X	X	X	X	X	X	X	X
Use safety clamps on truck tailgates	X									X
Maintain road markings and trench barriers	X									X
Maintain fencing								X		X
Apply insecticide					X				X	
Maintain buffer areas and grass		X	X	X		X	X	X		
Proper equipment maintenance	X						X			X
Spray water/oil/liquid asphalt			X	X						
Truck wash pad (to clean trucks)			X	X		X				
Maintain grass waterways, diversion ditches, rip rap		X	X					X		
Final grading of disturbed areas		X						X		
Revegetation of disturbed areas		X	X	X				X		
Chemical masking agent						X				
Lime on site	X				X	X			X	X
Workers supplied with aerators				X		X			X	X
Cover sludge daily					X	X		X	X	X
Water diverted away from site		X	X							

Source: US EPA.

13. SITE CLOSURE

In closing a biosolids landfill site, certain criteria must be met to make the site publicly acceptable. These criteria are established according to the type of landfill and the location, size, and ultimate use of the site. The procedures for site closure should be included in the operations manual and updated or modified as the original landfill plan is altered.

13.1. Ultimate Use

The ultimate use of the site should be described and illustrated in the O&M manual or in a separate document describing the closure of the site. The actual work involved in completing the site will depend on its ultimate use and on the care taken in day-to-day fill operations.

13.2. Grading at Completion of Filling

When each section of the landfill is completed, the final cover should be graded according to a predetermined plan. It is imperative that no biosolids become or remain exposed after the grading has been completed (50,51). Final grading of the site is to be performed after sufficient time has elapsed to allow for initial settlement. The final grading plan should be designed in accordance with the intended ultimate use of the landfill site. It is important that all biosolids be completely covered to the specified depth with cover material.

13.3. Landscaping

The landscaping plan should reflect the intended ultimate use of the landfill site. Where practical, landscaping may be done on completed sections before the entire fill project is completed (52).

13.4. Continued Leachate and Gas Control

Since decomposition of the organics in the biosolids may continue even after the landfill has been completed, an ongoing monitoring and control program must be maintained. Leachate and gas must be controlled even after the filling operations have stopped. The completion plans should clearly outline this program.

14. COSTS OF BIOSOLIDS DISPOSAL ON LAND (LANDFILL)

14.1. General

a. Most biosolids disposal systems have at least four definable components: storage, collection, haul, and disposal.
b. Treatment of biosolids is related to reducing the volume to a minimum before transporting. Typical unit processes used for volume reduction may include digestion, centrifugation, vacuum filtration, and drying beds (12–15,53). Costs associated with these processes are not considered to be part of biosolids hauling or landfilling but are very important in the overall biosolids handling train.
c. Storage cost is site-specific and depend largely upon the method selected in the biosolids handling train (44,45). It may be simply the costs associated with the purchase of bins for storage of secondary or primary biosolids, a dump truck for storage of digested biosolids that have been centrifuged or vacuum filtered, or the cost associated with drying beds.

d. Collection costs are dependent upon a time-labor relationship to transfer the biosolids from storage to the transporting vehicle, as a dump or tank truck (54). There may not be a collection cost associated with labor; however, a cost would be incurred to provide a vehicle during the loading period. Larger facilities may require that a driver be assigned to the vehicle during loading period. Collection cost might be significant when it is necessary to shovel biosolids from drying beds into trucks for transportation to the landfill. Collection costs are site and system specific.
e. Transportation costs are associated with such parameters as truck cost, truck size, haul time, labor, and operating cost per unit time for items such as depreciation, fuel, insurance, maintenance, and so on (44). Operating cost may be estimated from manufacturer's rating information and used in conjunction with estimates of biosolids production from various wastewater treatment processes.
f. Disposal costs are related to the operation and management of the final disposal facility. This cost should be minimal if the facility will integrate ultimate biosolids disposal with the disposal of refuse. When this is possible, the disposal costs may only include the costs of unloading and a landfill fee. On the other hand, if the landfill is to receive only waste biosolids; costs may be very significant as other equipment for operation of the landfill will be required. The equipment used for landfill operation may include units for excavation, placing, covering, and compaction of fill.
g. The lowest possible moisture content attainable at a reasonable cost should be produced for economical biosolids hauling and landfill operations. A reduction of moisture content will produce a savings in storage, initial equipment, operating, and labor cost.

14.2. Hauling of Biosolids

14.2.1. Required Input Data

a. Average wastewater flow (MGD).
b. Biosolids volume (gal/MG).
c. Raw biosolids concentration (%).
d. Dewatered biosolids concentration (%).
e. Vehicle loading time (h).
f. Round-trip haul time (h).
g. Vehicle capacity (yd^3).
h. Solids capture in dewatering process (%).
i. Distance to disposal site (mile).

14.2.2. Design Parameters

a. Biosolids volume/MG treated (Table 13).
b. Biosolids concentration (1.5–15%) (Table 13).
c. Cake concentration (6–60%) (Table 14).
d. Vehicle capacity (yd^3/truck).
e. Truck loading time (0.5–2 h).
f. Haul time, local conditions (h).
g. Daily work schedule (6–8 h).
h. Solids capture (70–99%) (Table 14).

14.2.3. Design Procedure

a. Compute the biosolids volume hauled yd^3/d.

$$V_B = \frac{(Q)\,(\mathrm{BF})\,(\mathrm{SS})}{(\mathrm{CSS})\,(7.48)\,(27)} \tag{1}$$

where V_B is the volume of biosolids (yd^3/d); Q is the wastewater flow (MGD); BF is the Biosolids flow (gal/MG) (Table 13); SS is the suspended solids in biosolids flow (%) (Table 13); CSS is the cake suspended solids (%) (14).

Table 13
Normal Quantities of Biosolids Produced by Various Treatment Processes

Wastewater treatment process	Gallons sludge/ MG treated	Solids (%)	Sludge specific gravity
Primary sedimentation			
Undigested	2950	5	1.02
Digested in separate tanks	1450	6	1.03
Trickling filter	745	7.5	1.025
Chemical precipitation	5120	7.5	1.03
Primary sedimentation and activated sludge			
Undigested	6900	4	1.02
Digested in separate tanks	2700	6	1.03
Activated sludge			
Waste sludge	19,400	1.5	1.005
Septic tanks, digested	900	10	1.04
Imhoff tanks, digested	500	15	1.04

Source: US EPA.

Table 14
Process Efficiencies for Dewatering of Wastewater Biosolids

Unit process	Solids capture (%)	Cake solids (%)
Centrifugation		
Solid bowl	80–90	5–13
Disc-nozzle	80–97	5–7
Basket	70–90	9–10
Dissolved air flotation	95	4–6
Drying beds	85–99	8–25
Filter press	99	40–60
Gravity thickener	90–95	5–12
Vacuum filter	90+	28–35

Source: US EPA.

b. Calculate the number of vehicles for collection and hauling of the biosolids.

$$N = \frac{(V_B)\ (\mathrm{LT} + \mathrm{HT})}{(\mathrm{HPD})\ (\mathrm{CAP})} \tag{2}$$

where N is the number of trucks required; LT is the loading time (h) (0.5–2 h); HT is the round-trip haul time (h) (local conditions); HPD is the work schedule (h/d) (6–8); CAP is the vehicle capacity (yd^3/truck) (3–12).

c. Compute the tons of biosolids hauled per day.

$$\mathrm{TBH} = \frac{(Q)\ (\mathrm{BF})\ (\mathrm{SS})\ (\mathrm{SCAP})\ (8.34)}{(100)\ (\mathrm{CSS})\ (2000)} \tag{3}$$

where TBH is the tons of biosolids hauled per day (t/d); SCAP is the solids capture (%) (Table 14).

14.2.4. Output Data

a. Volume of biosolids to be dewatered (gpd).
b. Initial moisture content (%).

Table 15
Fuel Energy Consumption Rates for Some Typical Construction Equipment

Equipment	Average diesel fuel (gal/h)	Equipment	Average diesel fuel (gal/h)
Caterpillar D-6	5.2	Grader-25,000 lb	4.4
Caterpillar D-8	10.8	28,000 lb	4.8
Excavator-0.5 yd^3	3.4	30,000 lb	5.2
1 yd^3	5	40,000 lb	6
1.25–1.5 yd^3	8.8	54,000 lb	7.9
1.5–2 yd^3	11.1	Track loader-1 yd^3	2.4
Wheel loader 1.5 yd^3	3	1.5 yd^3	3.4
2 yd^3	3.7	2 yd^3	4.2
3 yd^3	4.6	2.5 yd^3	5.7
4 yd^3	6.2	3 yd^3	7.4
5 yd^3	9	4 yd^3	11.3
7 yd^3	13.2	Tractor-scraper, small	4.9
		medium	11.4
		large	15.8

Source: US EPA.

c. Final moisture content (%).
d. Volume of biosolids hauled (yd^3/d).
e. Truck capacity (yd^3).
f. Time to make one load (h).
g. Work schedule (h/d).
h. Number of trucks required.
i. Tons of biosolids hauled per day (t/d).
j. Distance to disposal site (mile).

14.3. Energy Requirements

Actual fuel consumption varies considerably with specific biosolids, site and operating conditions (32). Fuel consumption rates for some typical construction equipment performing light to medium work are given in Table 15. One case study that used biosolids, landfill operation was estimated to consume 700,000 Btu/d/t of biosolids (1 gal diesel fuel is equivalent to 140,000 BTU).

14.4. Costs

Construction and O&M costs for the two trench and three area fill methods are shown in Figs. 1–4. All costs are in 1978 USD (Cost Index = 235.78). To obtain the values in terms of the present 2006 USD using the Cost Index for Utilities shown in Appendix, multiply the costs by a factor of 2.24 (29). Also, take notice of the following items (32):

a. Site and equipment costs include land (2500 USD/acre), site preparation (clearing, grubbing, surface water control ditches and ponds, monitoring wells, soil stockpiles, roads, and facilities), equipment purchase, and engineering (6%). Actual fill area consumes 50% of total site area.
b. Operating costs include labor [8 USD/h (2006 USD 18), including fringe, overhead, administration], equipment fuel, maintenance and parts; utilities; laboratory analysis of water samples; supplies and materials.
c. Actual costs vary considerably with specific biosolids and site conditions (28,32,55).

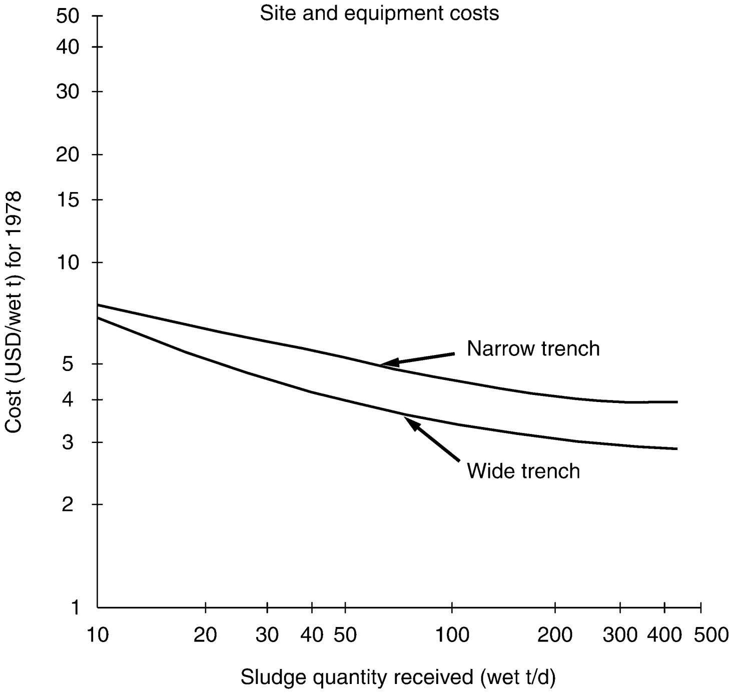

Fig. 1. Site and Equipment Costs for Narrow and Wide Trench Landfill (*Source:* US EPA).

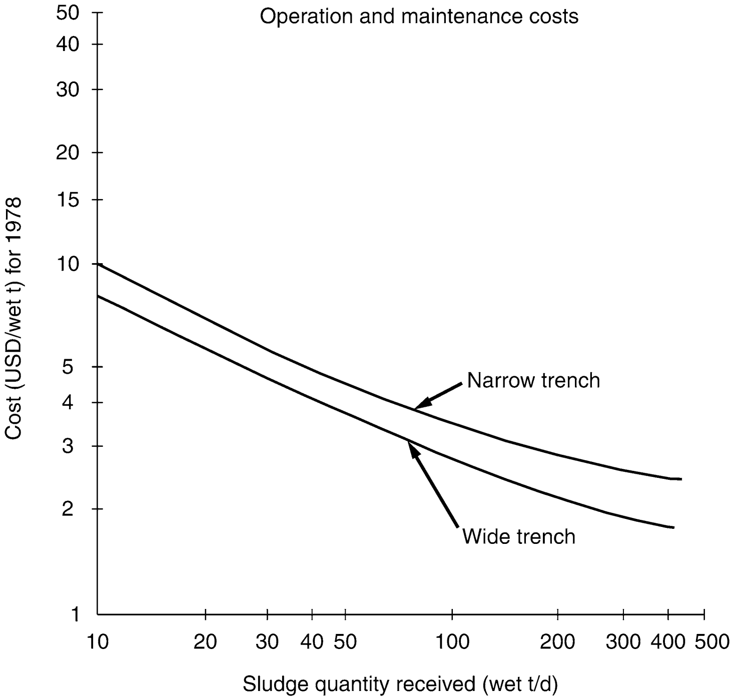

Fig. 2. Operation and Maintenance Costs for Narrow and Wide Trench Landfill (*Source:* US EPA).

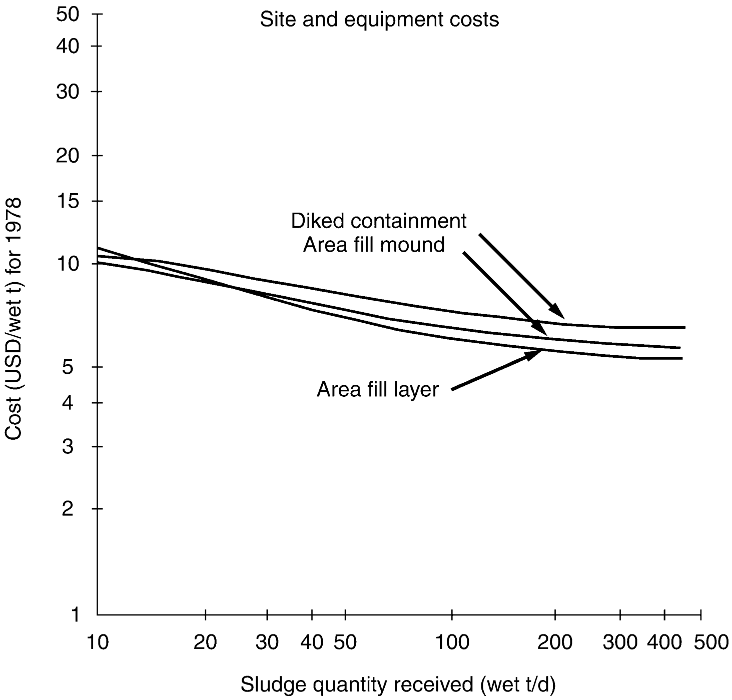

Fig. 3. Site and Equipment Costs for the Three Types of Area Landfill (*Source:* US EPA).

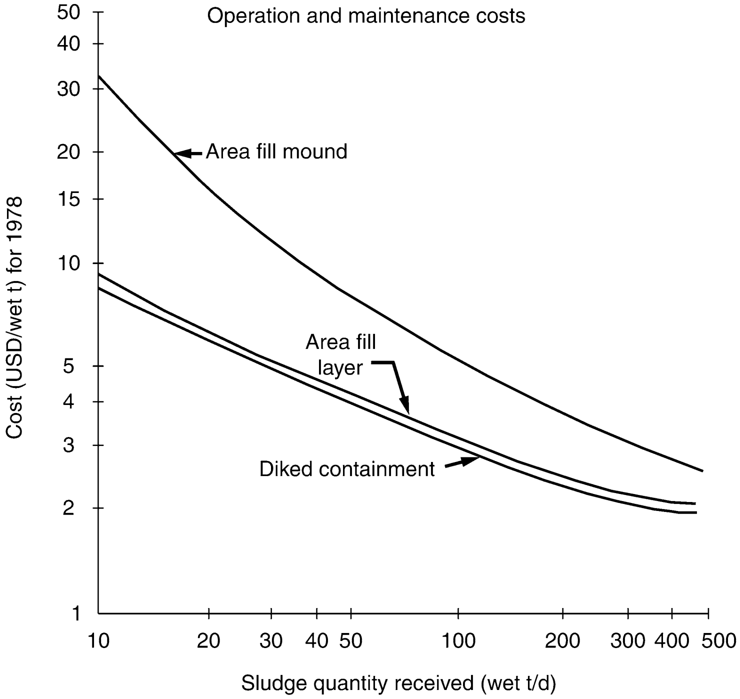

Fig. 4. Operation and Maintenance Costs for the Three Types of Area Landfill (*Source:* US EPA).

15. EXAMPLES

15.1. Example 1 Typical Biosolids Application Rate Scenario

The recommended minimum amount of nitrogen needed by a typical corn crop to be grown in New Jersey is 120 lb/acre/yr. Biosolids containing 3% nitrogen could be applied at up to 5.4 dry t/acre if used to supply all the nitrogen needed by the crop (i.e., no other nitrogen fertilizers used). A city producing 10 t/d of dry biosolids would require access to almost 700 acres of corn. If the biosolids contained only 1.5% nitrogen, twice as many tons could be applied per acre, requiring only half as many acres to land apply the same amount of biosolids generated.

15.2. Example 2 Hauling of Biosolids

A 1 MGD wastewater treatment plant is planned to dispose of its biosolids in a landfill. The plan had the following design parameters:

a. Biosolids production volume = 2700 gal/MG of wastewater flow.
b. Biosolids suspended solids content = 6%.
c. Biosolids cake solids content = 12%.
d. Solids capture = 95%.
e. Hauling truck capacity = 4 yd^3.
f. Truck loading time = 1.5 h.
g. Transport driving time to landfill = 2.5 h.

Compute the following:

a. Volume of biosolids to be hauled.
b. Number of vehicles required.
c. Tons of biosolids hauled per day.

Solution

a. Volume of biosolids to be hauled.

$$V_B = \frac{(Q)(BF)(SS)}{(\text{CSS})\,(7.48)\,(27)} \tag{1}$$

where V_B is the biosolids volume hauled (yd^3/d); Q is the wastewater flow (1 MGD); BF is the biosolids flow (2700 gal/MG); SS is the suspended solids in biosolids flow (6%); CSS is the cake suspended solids (12%).

$$V_B = \frac{(1)\,(2700)\,(6)}{(12)\,(7.48)\,(27)}$$

$$V_B = 6.7 \text{ yd}^3/\text{d}.$$

b. Number of vehicles required.

$$N = \frac{(V_B)\,(LT + HT)}{(\text{HPD})\,(\text{CAP})} \tag{2}$$

where N is the number of trucks required; V_B is the biosolids volume hauled (6.7 yd^3/d); LT is the loading time (1.5 h); HT is the round-trip haul time (2.5 h); HPD is the work schedule (8 h/d); CAP is the vehicle capacity (4 yd^3/truck).

$$N = \frac{6.7\ (1.5 + 2.5)}{(8)\,(4)}$$

$$N = 0.84.$$

Therefore use 1 truck at 2 trips/d.

c. Tons of biosolids hauled per day.

$$\text{TBH} = \frac{(Q)\ (\text{BF})\ (\text{SS})\ (\text{SCAP})\ (8.34)}{(100)\ (\text{CSS})\ (2000)} \tag{3}$$

where TBH is the tons of biosolids hauled per day (t/d); Q is the wastewater flow (1 MGD); BF is the biosolids flow (2700 gal/MG); SS is the suspended solids in biosolids flow (6%); SCAP is the solids capture (95%); CSS is the cake suspended solids (12%).

$$\text{TBH} = \frac{(1)\ (2700)\ (6)\ (95)\ (8.34)}{(100)\ (12)\ (2000).}$$

$$\text{TBH} = 5.3\ \text{t/d.}$$

NOMENCLATURE

BF	Biosolids flow (gal/MG)
CAP	Vehicle capacity (yd^3/truck)
CSS	Cake suspended solids (%)
HPD	Work schedule (h/d)
HT	Round-trip haul time (h)
LT	Loading time (h)
N	Number of trucks required
Q	Wastewater flow (MGD)
SCAP	Solids capture (%)
SS	Suspended solids in biosolids flow (%)
TBH	Tons of biosolids hauled per day (t/d)
V_B	Volume of biosolids (yd^3/d)

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APPENDIX

United States Yearly Average Cost Index for Utilities[a]

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

[a] From ref. 29.

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 55-page 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.

I. 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	30577	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	43560	square feet
acres	4047	square meters
acres	1.562×10^{-3}	square miles
acres	4840	square yards
acre-feet	43560	cubic feet
acre-feet	1233.5	cubic meters
acre-feet	325850	gallons (U.S.)
amperes (abs)	0.1	abamperes
amperes (abs)	1.036×10^{-5}	faradays/second
amperes (abs)	2.9980×10^{9}	statamperes

Multiply	by	to obtain
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	1,550.0	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
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 (Mu)
ares	0.02471	acre (US)
ares	1076	square feet
ares	100	square meters
ares	119.60	sq. yards
assay tons	29.17	grams

Multiply	by	to obtain
astronomical unit	1.495×10^{8}	kilometers
atmospheres (atm)	.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	in. of mercury (at 0°C)
atmospheres	1.033228	kgs/sq cm
atmospheres	10,332	kgs/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)

Multiply	by	to obtain
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.)
bars	0.98692	atmospheres
bars	10^6	dynes/sq cm
bars	1.0197×10^4	kgs/sq meter
bars	1000	millibar
bars	750.06	mm of Hg (0^oC)
bars	2,089	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. x 1 in.	cubic inches
boiler horsepower	33475	Btu (mean)/hour
boiler horsepower	34.5	pounds of water evaporated from and at 212^oF (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

Multiply	by	to obtain
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 x 10^{-4}	horsepower-hrs (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 x 10^{-4}	gram-calorie/second (cm^2)°C
Btu (mean)/hour(feet^2)°F	3.94 x 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 x 10^{-4}	watts/(cm^2)°C
Btu (mean)/hour(feet^2)°F	2.035 x 10^{-3}	watts/(in.^{2})°C
Btu(mean)/(hour)(feet^2) (°F/inch)	3.4448 x 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.)
Btu(mean)/(hour)(feet^2) (°F/inch)	1.442 x 10^{-3}	joules (abs)/(sec)(cm^2) (°C/cm)
Btu(mean)/(hour)(feet^2) (°F/inch)	1.442 x 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 x 10^4	cubic cm
buckets (British, dry)	4	gallons (British)
bushels (British)	1.03205	bushels (U.S.)

Multiply	by	to obtain
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 (1)
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-^{o}C	4186.8	joule/kg-^{o}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
candles(int)/square centimeter	3.1416	lamberts

Multiply	by	to obtain
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	kgs./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

Multiply	by	to obtain
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
centimeters/second	0.0328	fps
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.
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)

Multiply	by	to obtain
chu/(hr)(ft^2)(^{o}C/in.)	1	Btu/(hr)(ft^2)(^{o}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' x 4' x 4'	cubic feet
cords	128	cubic feet
cords	3.625	cubic meters
cord-feet	4' x 4' x 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	1,550	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.)
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.)

Multiply	by	to obtain
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

Multiply	by	to obtain
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
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}	gal (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	61023	cubic inches (U.S.)
cubic meters	1.308	cubic yards (U.S.)
cubic meters	264.17	gallons (U.S.)
cubic meters	999.973	liters
cubic meters	2113	pints (liq.)

Multiply	by	to obtain
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	gal (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
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)

Multiply	by	to obtain
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	$^{o}F=(^{o}C\times 9/5)+32$	fahrenheit
degree Celsius	$^{o}K=^{o}C+273.15$	kelvin
degree Fahrenheit	$^{o}C=(^{o}F-32)\times 5/9$	celsius
degree Fahrenheit	$^{o}K=(^{o}F+459.67)/1.8$	kelvin
degree Rankine	$^{o}K=^{o}R/1.8$	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)

Multiply	by	to obtain
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
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

Multiply	by	to obtain
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
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	96500	coulombs (abs)
faradays/second	96500	amperes (abs)

Multiply	by	to obtain
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, $60^{o}F$)	5.30×10^{-4}	pounds/square inch
feet of water	0.02950	atm
feet of water	0.8826	in. of mercury
feet of water at $39.2^{o}F$	0.030479	kilograms/square centimeter
feet of water at 39.2 ^{o}F	2988.98	newton/meter^{2} (N/m^{2})
feet of water at $39.2^{o}F$	304.79	kilograms/square meter
feet of water	62.43	pounds/square feet (psf)
feet of water at $39.2^{o}F$	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	kms/hr
feet/min	0.3048	meters/min
feet/min	0.01136	miles/hr
feet/sec	30.48	cms/sec
feet/sec	1.097	kms/hr

Multiply	by	to obtain
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	cms/sec/sec
feet/sec/sec	1.097	kms/hr/sec
feet/sec/sec	0.3048	meters/sec/sec
feet/sec/sec	0.6818	miles/hr/sec
feet/100 feet	1.0	per cent 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

Multiply	by	to obtain
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.)
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

Multiply	by	to obtain
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	8.0208	cfh
gpm	2.228×10^{-3}	cfs
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 m
gpm/sq ft	0.679	liter/sec/sq. meter

Multiply	by	to obtain
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	.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)

Multiply	by	to obtain
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
grains/imp.gal	14.254	parts/million
grains/US gal	17.118	mg/l
grains/US gal	17.118	parts/million
grains/US 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-hrs
gram-calories	1.1630×10^{-6}	kilowatt-hrs
gram-calories	1.1630×10^{-3}	watt-hrs
gram-calories	3.968×10^{-3}	British Thermal Units (Btu)

Multiply	by	to obtain
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
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.)

Multiply	by	to obtain
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/1,000 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

Multiply	by	to obtain
hemispheres	0.5	spheres
hemispheres	4	spherical right angles
hemispheres	6.2832	steradians
henries (abs)	10^9	abhenries
henries	1,000.0	millihenries
henries (abs)	1.1126×10^{-12}	stathenries
hogsheads (British)	63	gallons (British)
hogsheads (British)	10.114	cubic ft
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

Multiply	by	to obtain
horsepower-hours	641.7	kilogram-calories
horsepower-hours	2.737×10^5	kilogram-meters
horsepower-hours	0.7457	kilowatt-hours (abs)
horsepower-hours	26494	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)
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

Multiply	by	to obtain
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
joule-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

Multiply	by	to obtain
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	15432	grains
kilograms	1,000	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

Multiply	by	to obtain
kilogram-calories	3086	foot-pounds
kilogram-calories	1.558×10^{-3}	horsepower-hours
kilogram-calories	4,186	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
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)

Multiply	by	to obtain
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	mms. of mercury
kilograms/sq cm	2,048	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
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

Multiply	by	to obtain
kilograms/sq meter	0.00142234	pounds/square inch
kilograms/sq mm.	10^6	kgs./square meter
kilojoule	0.947	Btu
kilojoules/kilogram	0.4295	Btu/pound
kilolines	1,000.0	maxwells
kiloliters	10^3	liters
kilometers	10^5	centimeters
kilometers	3281	ft
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

Multiply	by	to obtain
kilowatts	14.34	kg-cal/min
kilowatts	10^3	watts
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	6,080	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	joule (abs)/cm^2
leagues (nautical)	3	miles (nautical)
leagues (statute)	3	miles (statute)
light years	63274	astronomical units
light years	9.4599×10^{12}	kilometers
light years	5.8781×10^{12}	miles

Multiply	by	to obtain
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.)
liters	0.02838	bushels (U.S. dry)
liters	1000.028	cubic centimeters
liters	0.035316	cubic feet
liters	61.025	cu in
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	16894	minims (Br.)
liters	16231	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

Multiply	by	to obtain
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
lux	1	lumens/sq meter
lux	10^{-4}	phots
maxwells	0.001	kilolines
maxwells	10^{-8}	webers

Multiply	by	to obtain
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

Multiply	by	to obtain
meters/second	2.23693	miles/hour
meters/second	0.03728	miles/minute
meters/sec/sec	100.0	cms/sec/sec
meters/sec/sec	3.281	feet/sec/sec
meters/sec/sec	3.6	kms./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
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	0.8347	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

Multiply	by	to obtain
microns	10^{-6}	meters
miles (naut.)	6,080.27	feet
miles (maut.)	1.853	kilometers
miles (naut.)	1.853	meters
miles (naut.)	1.1516	miles (statute)
miles (naut.)	2,027	yards
miles (statute)	1.609×10^5	centimeters
miles (statute)	5,280	feet
miles (statute)	6.336×10^4	inches
miles (statute)	1.609	kilometers
miles (statute)	1,609	meters
miles (statute)	0.8684	miles (naut.)
miles (statute)	320	rods
miles (statute)	1,760	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	kms/min
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	kms/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

Multiply	by	to obtain
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/US 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
milligrams/sq cm	10^{4}	milligrams/sq meter

Multiply	by	to obtain
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
milliters	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}	kgs/sq cm

Multiply	by	to obtain
millimeter of mercury (0^oC)	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	3,785.0	cubic meters
million gallons	3.785	mega liters (1 x 10^6)
million gallons/day (mgd)	1.547	cu ft/sec
mgd	3,785	cu m/day
mgd	0.0438	cubic meters/sec
mgd	43.808	liters/sec
mgd/acre	9,360	cu m/day/ha
mgd/acre	0.039	cu meters/hour/sq meter
mils	0.002540	centimeters
mils	8.333 x 10^{-5}	feet
mils	0.001	inches
mils	2.540 x 10^{-8}	kilometers
mils	25.40	microns
mils	2.778 x 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 (Ida., 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 x 10^{-4}	quadrants
minutes (angles)	2.909 x 10^{-4}	radians
minutes (angle)	60	seconds (angle)
months (mean calendar)	30.4202	days

Multiply	by	to obtain
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.
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

Multiply	by	to obtain
OHM (International)	1.0005	OHM (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}	cu ft
ounces (Br. fl)	1.73457	cu in
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)

Multiply	by	to obtain
ounces (troy)	0.08333	pounds (troy)
ounces (troy)	3.061×10^{-5}	tons (long)
ounces (troy)	3.429×10^{-5}	tons (short)
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}	cu ft
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 gal
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

Multiply	by	to obtain
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
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	cu feet
pints (liq.)	4.732×10^{-4}	cu meters
pints (liq.)	6.189×10^{-4}	cu yards
pints (liq.)	0.125	gallons
pints (liq.)	0.4732	liters
pints (liq.)	16	ounces (U.S. fluid)

Multiply	by	to obtain
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)
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)

Multiply	by	to obtain
pounds (troy)	5,760	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	14544	Btu (mean)
pounds of water	0.0160	cu ft
pounds of water	27.68	cu in
pounds of water	0.1198	gal
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	kgs.-cms. squared
pounds-feet squared	144	pounds-ins. squared
pounds-inches squared	2,926	kgs.-cms. squared
pounds-inches squared	6.945×10^{-3}	pounds-feet squared
pounds/acre	0.0104	grams/sq ft

Multiply	by	to obtain
pounds/acre	0.1121	grams/sq meter
pounds/acre	1.121	kg/ha
pounds/acre	112.1	kilograms/sq km
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

Multiply	by	to obtain
pounds/gal	454 g/3.785 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
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

Multiply	by	to obtain
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	cu cms
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}	cu 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
radians/sec/sec	573.0	revs/min/min
radians/sec/sec	9.549	revs/min/sec
radians/sec/sec	0.1592	revs/sec/sec

Multiply	by	to obtain
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

Multiply	by	to obtain
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
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}	sq 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	sq in
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

Multiply	by	to obtain
square foot-square foot (moment of area)	20736	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 cms.-cms sqd.
square inches-inches sqd.	4.823×10^{-5}	sq feet-feet sqd.
square kilometers	247.1	acres
square kilometers	10^{10}	sq cms
square kilometers	10.76×10^6	sq ft
square kilometers	1.550×10^9	sq inches
square kilometers	10^6	square meters
square kilometers	0.3861006	square miles (U.S.)
square kilometers	1.196×10^6	sq yd
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	sq cms
square meters	10.76387	square feet (U.S.)
square meters	1550	square inches
square meters	3.8610×10^{-7}	square miles (statute)
square meters	10^6	sq 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

Multiply	by	to obtain
square millimeters	1.973×10^{3}	circular mils
square millimeters	0.01	square centimeters
square millimeters	1.076×10^{-5}	sq 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	8,361	sq cms
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}	sq 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

Multiply	by	to obtain
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.)
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	kgs./sq meter
tons (short)/sq ft	13.89	pounds/sq inch

Multiply	by	to obtain
tons (short)/sq in	1.406×10^6	kgs/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, $0^\circ C$)	133.322	newton/meter^2
townships (U.S.)	23040	acres
townships (U.S.)	36	square miles
tuns	252	gallons
volts (abs)	10^8	abvolts
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

Multiply	by	to obtain
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 (internationl)	1.0002	watt (absolute)
watt/(cm^2)(^{o}C/cm)	693.6	Btu/(hr)(ft^2)(^{o}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
weeks	604,800	seconds

Multiply	by	to obtain
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)

II. 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 caesium-133 atom when unperturbed by exterior fields.

An ampere is the constant current which if maintained in two parallel recti-linear 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 (^{o}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 cent of a circle by an arc of the circle equal in length to the radius of the circle.

A steradian is the solid angle which, 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.

III. 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 1.000 028 dm^3. This 1901 definition applied for the purpose of the 1963 Weights and Measures Acts.

By a resolution of the twelfth Conference General des Poids et Mesures (CGPM) in 1964 the word "liter" is now recognised as a special name for the dm^3, but is not used to express high precision measurements. It will be 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 that force which, 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 meter.

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 which 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: $du = c_v\ dT$, 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 $J/^{o}K$.

The Specific Entropy (s) of a substance is the entropy per unit mass, i.e. $J/kg^{o}K$.

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.

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.

IV. PHYSICAL CONSTANTS

Standard Temperature and Pressure (S.T.P.) = 273.15 °K and 1.013 x 10^{5} N/m^{2}
= 0 °C and 1.013 bar
= 0 °C and 760 mm Hg

Molecular Volume of ideal gas at S.T.P. = 22.41 liters/mol

Gas Constant (R) = 8.314 J/mol °K

$RT_{(273.15\ ^{o}K)}$ = 2.271 x 10^{3} J/mol

Avogadro Constant = 6.023 x 10^{23}/mol

Boltzmann Constant = 1.3805 x 10^{-23} J/K

Faraday Constant = 9.6487 x 10^{4} °C/mol (=A s/mol)

Planck Constant = 6.626 x 10^{-34} J sec

Stefan-Boltzman Constant = 5.6697 x 10^{-8} W/m^{2} K^{4}

Ice Point of Water = 273.15 °K (0°C)

Triple Point of Water = 273.16 °K (0.01 °C)

Speed of light = 2.998 x 10^{8} m/sec

Acceleration of Gravity (Standard) = 9.80665 m/s^{2} [Take g as 9.81 m/s^{2}]
(Greenwich) = 9.81188 m/s^{2}

Universal Constant of Gravitation = 6.670 x 10^{-11} Newton m^{2}/kg^{2}

Mass of hydrogen atom = 1.6734 x 10^{-27} kg

V. 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,[a] $\sigma \times 10^3$ (lb/ft)	Vapor Pressure, p_v (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

VI. PERIODIC TABLE OF THE ELEMENTS

Complements of

Lenox Institute of Water Technology

Groups ⇨ Periods & sub-shells ⇩	1 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 1s	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 Carbon	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 **Al** 26.9815 Aluminum	14 **Si** 28.0855 Silicon	15 **P** 30.9738 Phosphorus	16 **S** 32.066 Sulfur	17 **Cl** 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** 72.561 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 **Tl** 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) Rutherfordium	105 Ha (262) Hahnium	106 Sg (263) Seaborgium	107 Ns (262) Nielsbohrium	108 Hs (265) Hassium	109 Mt (266) Metnenium	110		112						

	19	20	21	22	23	24	25	26	27	28	29	30	31	32
6 4f	58 **Ce** 140.115 Cerium	59 **Pr** 140.91 Praseodymium	60 **Nd** 144.24 Neodymium	61 Pm (145) Promethium	62 **Sm** 150.35 Samarium	63 **Eu** 167.26 Europium	64 **Gd** 157.25 Gadolinium	65 **Tb** 158.925 Terbium	66 **Dy** 162.50 Dysprosium	67 **Ho** 164.930 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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