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Technology Transfer

# Sludge Treatment and Disposal

EPA-625/4-78-012

## Sludge Treatment

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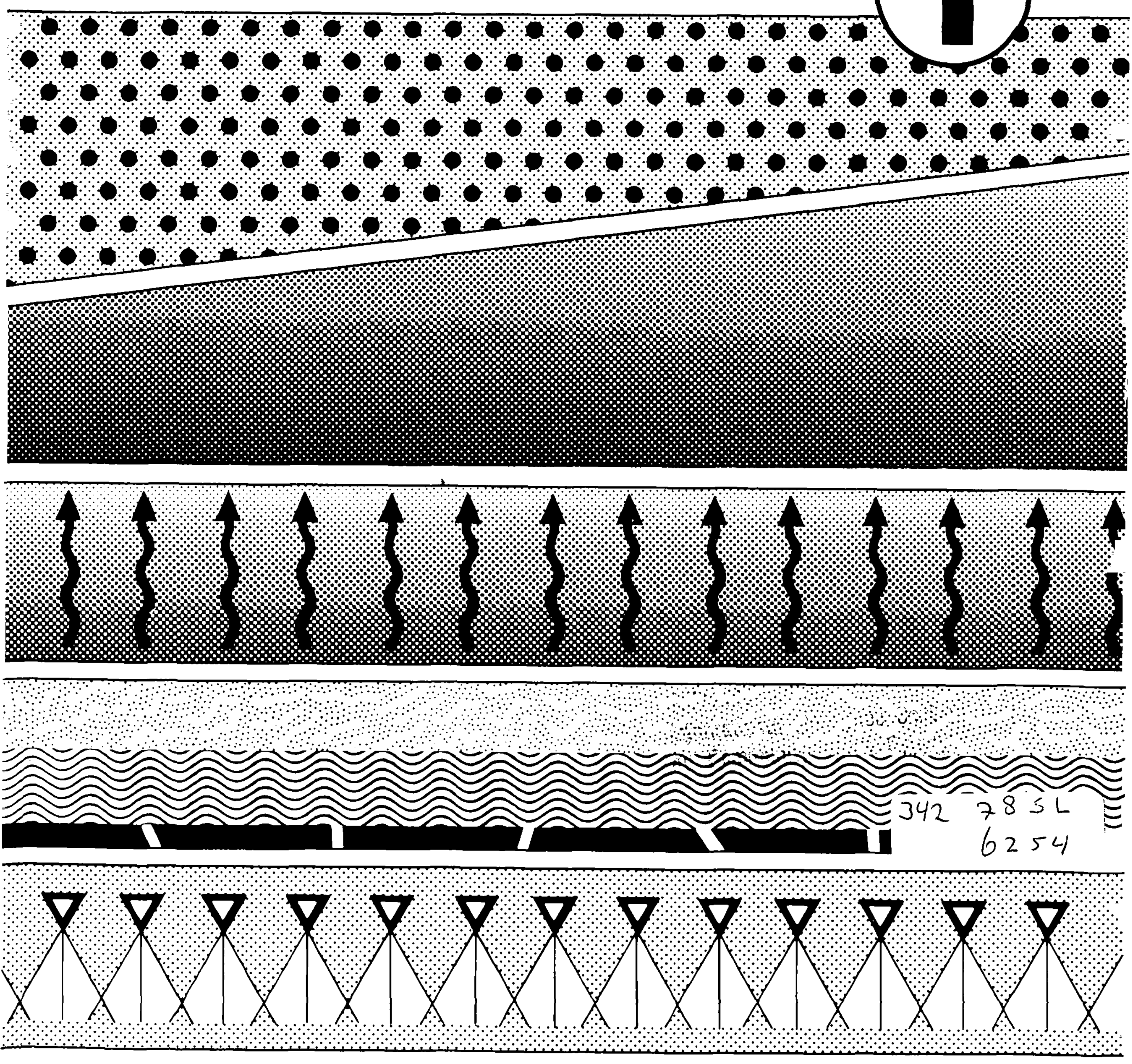
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# Sludge Treatment and Disposal

## Sludge Treatment

## Volume 1

Environmental Research Information Center
Cincinnati, Ohio 45268

## NOTICE

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# Acknowledgments

This seminar publication contains material prepared for the U.S. Environmental Protection Agency Technology Transfer Program. It has been presented at Technology Transfer Design Seminars held at various locations throughout the United States. The information in this publication was prepared by the following:

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Anaerobic Digestion of Municipal Wastewater Sludges. N. A. Mignone, Envirex Inc., Waukesha, Wis.

Aerobic Digestion of Municipal Wastewater Sludges. N. A. Mignone, Envirex Inc., Waukesha, Wis.

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# Contents

# Introduction

Of the utilization and disposal options available for sludge, each has its own specific set of environmental problems. In order to implement any policy, resolution to a number of problems that presently inhibit sludge management must be developed. These problems can be summarized and categorized into four general areas:

- Public health issues.
- Technological factors.
- Intermedia issues.
- Social/economic/institutional factors.

This section briefly discusses the seminar publication's contents, their evolution, and the U.S. Environmental Protection Agency's (EPA) sludge management research and development program.

This publication is the culmination of two years of intensive work and eight seminars that were presented around the nation in an effort to develop and disseminate the latest information on the design of sludge treatment and disposal systems. It presents in great detail technical information for the following sludge treatment and disposal processes:

- Lime Stabilization.
- Anaerobic Digestion.
- Aerobic Digestion.
- Thermal Sludge Conditioning.
- Thickening.
- Dewatering.
- Incineration and Pyrolysis.
- Composting.
- Land Utilization.
- Landfilling.

The discussion of each process includes where possible a presentation of performance data for existing operations as well as operation and maintenance experiences and energy and cost information. Each chapter includes one or more design examples to illustrate step-by-step, the philosophy, rationale, and methodology behind the design of the particular process.

The "Lime Stabilization" discussion gives information for determining lime requirements as well as a detailed case history for a 1.0 Mgal/day (.04 $m^3/s$) plant. Comparative designs and cost information are presented for both the "Lime Stabilization" and "Anaerobic Digestion" processes in 4 (.18 $m^3/s$) and 40 (1.175 $m^3/s$) Mgal/day plants. A list of installations employing lime stabilization is included.

Both the "Anaerobic Digestion" and "Aerobic Digestion" chapters thoroughly review the pertinent parameters for such biological processes and include a design relationship between the percent reduction in volatile solids and sludge age and digestion temperature. The "Thickening" chapter provides design examples for two different plant sizes which include a detailed cost effective analysis for choosing the alternative techniques of gravity, dissolved air flotation, centrifugation, and no thickening. The "Dewatering" discussion includes various schemes for designing and operating drying beds, the continuous belt filter presses, and both the plate and frame and recessed chamber pressure filters. Standard as well as membrane and diaphragm pressure filters are discussed. The electric or infrared furnace is explained in the chapter on "Incineration and Pyrolysis," and performance data are presented. The plant scale partial pyrolysis (starved air combustion) work done at the Contra Costa County Sanitation District is described in detail. Considerable discussion is devoted to the use of alternative fuels and energy recovery.

The chapter on "Composting" discusses at length the work on forced aeration static pile composting at Beltsville, Md.; Bangor, Maine; and Durham, N.H. European developments with mechanical systems are also covered. Very detailed design information applicable to any size system is presented in both the "Land Utilization" and "Landfilling" chapters along with step-by-step examples.

EPA sludge management research and development program encompasses four major technical areas: processing and treatment, utilization, disposal, and health and ecological effects. The primary objective of the program is to develop new and improved technology and management schemes which will enable communities to solve problems associated with the residues or byproducts of wastewater treatment in a cost effective and environmentally acceptable manner.

The present state-of-the-art provides adequate (but expensive) capability to dewater sludges. Incineration practice is well established with exception of the potential impact of air emissions on health and ecology. However, coincineration (e.g., sludge plus solid waste) and pyrolysis technology is just emerging. Controversy continues both within and outside the Agency with regard to the environmental acceptability of applying municipal sludges to the land. This is especially true for agricultural uses. Heavy metals (especially cadmium), complex organics, and microbiological contaminants are the constituents of primary concern.

Specific examples of technological gaps that presently exist are:

- Cost of sludge processing and disposal is a *major* factor in wastewater treatment.
- Methods of converting sludge to beneficial byproducts are in the embryonic stages.
- Limited confidence exists in the efficacy of local industrial pretreatment programs for metals removal and methods for monitoring their effectiveness.
- Relative risks associated with land application need to be established with greater precision.
- Varying climatic and soil conditions as well as varying sludge composition require evaluation for a variety of sludges with optimum combinations of soil and vegetation.
- Methods for removing toxicants at the treatment plant are in the development stage; application is impeded because of economics of technology.

## PROCESSING AND TREATMENT

Sludge must undergo some processing or treatment to prepare it for ultimate disposition.

The goal of processing and treatment research and development is to produce technology alternatives which can be used to prepare the sludge for application to the land or for one of the conversion processes so that the total cost of handling or disposal is minimized.

Implementation of the program is focused on the following objectives:

- Evaluate the efficacy of pretreatment as an option to minimize toxicants in sludge.
- Characterize the nature of, and the dewatering properties of, "new" sludges using existing, upgraded and new technology.
- Develop hardware capable of producing a substantially drier sludge cake.
- Develop and define performance of existing and new processes for stabilizing sludge (anaerobic digestion, auto thermal thermophilic aerobic digestion, composting, etc.).
- Investigate ways to minimize energy consumption while simultaneously maximizing fuel production (activated carbon enhancement, solar heating, etc.).
- Determine cost and environmental impact of sludge processing systems.
- Provide guidance on technology for disinfection (up through sterilization) of sludge.

## CONVERSION PROCESSES

This part of the research program has been divided between efforts devoted to upgrading conventional incineration and tasks oriented toward development of new processes.

Current program objectives directed to meeting these needs include several projects, ongoing and planned to:

- Develop techniques for substitution of more abundant, less costly supplemental fuels such as coal and solid wastes (incineration and co-incineration).
- Develop processes and hardware for pyrolysis, co-pyrolysis and starved-air combustion.
- Characterize emissions to determine levels of potential pollutants (gaseous, liquid, solid) contained in emissions from sludge conversion facilities.
- Establish the "least cost" approaches to sludge conversion to the satisfaction of administrators, technologists and the general public.
- Evaluation of cementation processes and other beneficial use alternatives.

## LAND APPLICATION—MANAGEMENT

The objective relating to land application management is to develop methods and technology to control the transformation and/or movement of pollutants through the soil, plants, groundwater, and human food chain. The function of research and development associated with the health and ecological area is to analyze, evaluate, and interpret the data for purposes of establishing safe loading rates.

It is anticipated that accomplishment of the primary objectives will result in the establishment of management schemes for a variety of sludges with optimum combinations of soil and vegetation. Practices can then be defined for applying sludge to the land for purposes of reclaiming marginal or sub-marginal land, determining agricultural uses for both food and fiber, and landfill disposal.

## HEALTH EFFECTS

The difficulty in resolving this issue is that data which will permit a definitive evaluation and decision regarding the significance of sludge in the human food chain impact do not exist to the satisfaction of the several scientific disciplines involved. EPA is, therefore, working cooperatively with other Federal agencies, particularly USDA and FDA, to develop the information required to resolve the issue. Information developed by others, notably universities, State agencies, and municipalities also is being obtained.

Some current work directed to this issue includes:

- Evaluation of current knowledge of potential health effects.
- Determine viral contamination of ground and surface water of a land reclamation site.
- Developing methods for isolating viruses and chemicals.
- Characterize type, quantity and biological persistence of biologicals, trace metals, and other organic and inorganic substances in the environs of a sludge disposal site.
- Determine the potential of biologicals, metals, and organic substances entering the human food chain when digested sludge is used as a fertilizer.
- Study heavy metal uptake in beef animals grazed on sludge amended pasture.

Chapter 1

# Lime Stabilization of Wastewater Treatment Plant Sludges

## INTRODUCTION

Sludge constitutes the most significant byproduct of wastewater treatment; its treatment and disposal is perhaps the most complex problem which faces both the designer and operator. Raw sludge contains large quantities of microorganisms, mostly fecal in origin, many of which are pathogenic and potentially hazardous to humans. Sludge processing is further complicated by its variable properties and relatively low solids concentration. Solutions have long been sought for better stabilization and disposal methods which are reliable and economical and able to render sludge either inert or stable.

Lime stabilization has been shown to be an effective sludge disposal alternative when there is a need to:

A. Provide alternate means of sludge treatment during the period when existing sludge handling facilities, e.g., anaerobic or aerobic digesters, are out of service for cleaning or repair.
B. Supplement existing sludge handling facilities, e.g., anaerobic or aerobic digesters, incineration or heat treatment, due to the loss of fuel supplies or because of excess sludge quantities above design.
C. Upgrade existing facilities or construct new facilities to improve odor, bacterial, and pathogenic organism control.

Lime stabilization has been demonstrated to effectively eliminate odors. Regrowth of pathogens following lime stabilization is minimal. Of the organisms studied, only fecal streptococci have a potential for remaining viable.

Lime stabilized sludges are suitable for application to agricultural land; however, lime stabilized sludges have lower soluble phosphate, ammonia nitrogen, total Kjeldahl nitrogen, and total solids concentrations than anaerobically digested primary/waste activated mixtures from the same plant.

The purpose of this chapter is to present a review of stabilization and disinfection of municipal wastewater treatment plant sludges using lime stabilization, including specific design considerations. Two design examples incorporating lime stabilization into a 4 and 40 Mgal/d (0.18 and 1.75 $m^3/s$) wastewater treatment plant have been included to demonstrate the design procedure. Comparisons of the performance, capital and annual operation and maintenance costs for lime stabilization and anaerobic digestion were included for each design example. To further illustrate the application of lime stabilization techniques to small plants and/or facilities in need of an emergency sludge-handling process, an actual case history of lime stabilization at a 1 Mgal/d (0.04 $m^3/s$) facility was also included. The case history includes capital and annual operation and maintenance costs; chemical, bacterial, and pathological properties; and land application techniques.

## LIME STABILIZATION PROCESS DESCRIPTION

### Background

Historically, lime has been used to treat nuisance conditions resulting from open pit privies and from the graves of domestic animals. Prior to 1970. there was only a small amount of quantitative information available in the literature on the reaction of lime with sludge to make a more stable material. Since that time, the literature contains numerous references concerning the effectiveness of lime in reducing microbiological hazards in water and wastewater.[1-3] Information is also available on the bactericidal value of adding lime to sludge. A report of operations at the Allentown, Pa., wastewater treatment plants states that conditioning an anaerobically digested sludge with lime to pH 10.2 to 11, vacuum filtering and storing the cake destroyed all odors and pathogenic enteric bacteria.[4] Kampelmacher and Jansen[5] reported similar experiences. Evans[6] noted that lime addition to sludge released ammonia and destroyed bacillus coli and that the sludge cake was a good source of nitrogen and lime to the land.

Lime stabilization of raw sludges has been conducted in the laboratory and in full-scale plants. Farrel et al.[7] reported, among other results, that lime stabilization of primary sludges reduced bacterial hazard to a negligible value, improved vacuum filter performance, and provided a satisfactory means of stabilizing sludge prior to ultimate disposal.

Paulsrud and Eikum[8] reported on the effects of long-term storage of lime stabilized sludge. Their research included laboratory investigations of pH and microbial activity over periods up to 28 days.

Pilot scale work by C. A. Counts et al.[9] on lime stabilization showed significant reductions in pathogen populations and obnoxious odors when the sludge pH was greater than 12. Counts conducted growth studies on greenhouse and outdoor plots which indicated that the

Table 1-1.—Lime required for stabilization to pH 12 for 30 minutes

| Sludge type | Percent solids | Average lbs[a] $Ca(OH)_2$/lbs dry solids | Range lbs[a] $Ca(OH)_2$/lbs dry solids | Total[b] volume treated | Average total solids, mg/l | Average initial pH | Average final pH |
|---|---|---|---|---|---|---|---|
| Primary sludge[c] | 3–6 | 0.12 | 0.06–0.17 | 136,500 | 43,276 | 6.7 | 12.7 |
| Waste activated sludge | 1–1.5 | 0.30 | 0.21–0.43 | 42,000 | 13,143 | 7.1 | 12.6 |
| Septage | 1–4.5 | 0.20 | 0.09–0.51 | 27,500 | 27,494 | 7.3 | 12.7 |
| Anaerobic | 6–7 | 0.19 | 0.14–0.25 | 23,500 | 55,345 | 7.2 | 12.4 |

[a]Numerically equivalent to kg $Ca(OH)_2$ per kg dry solids.
[b]Multiply gallons × 3.785 to calculate liters.
[c]Includes some portion of waste activated sludge.

disposal of lime stabilized sludge on cropland would have no detrimental effects.

A research and demonstration contract was awarded to Burgess & Niple, Ltd. in March 1975 to complete the design, construction, and operation of full-scale lime stabilization facilities for a 1 Mgal/d (0.4 $m^3$/s) wastewater treatment plant, including land application of treated sludges. The contract also included funds for cleaning, rehabilitation, and operating an existing anaerobic sludge digester. Concurrent with the research and demonstration project, a considerable amount of full-scale lime stabilization work was completed by cities in Ohio and Con-

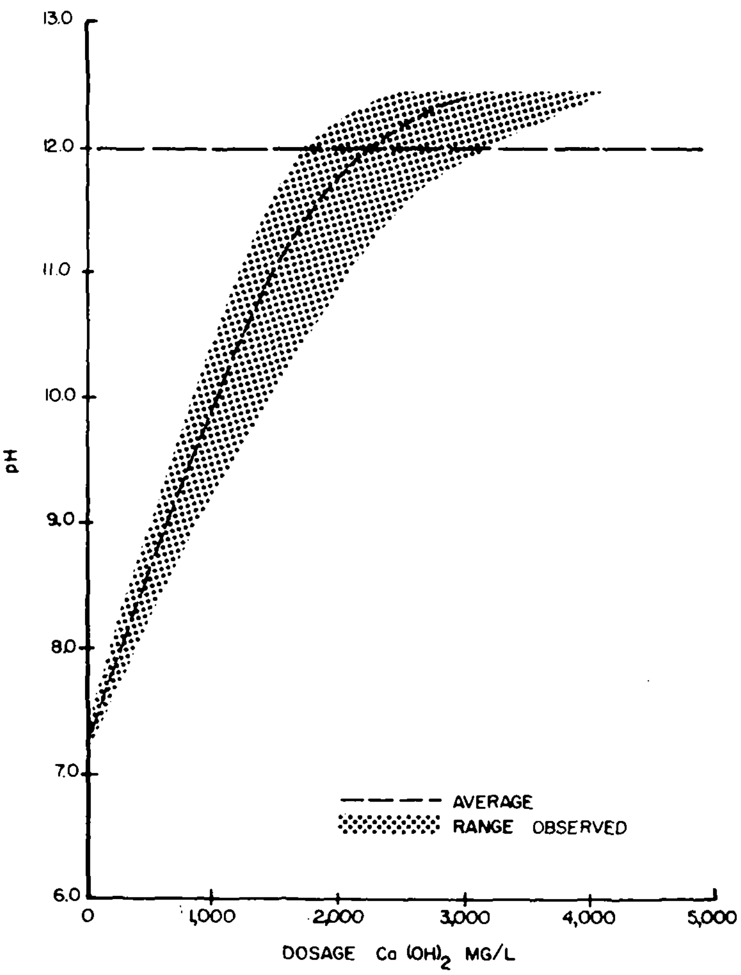



Figure 1-1.—Combined lime dosage versus pH for all sludges.

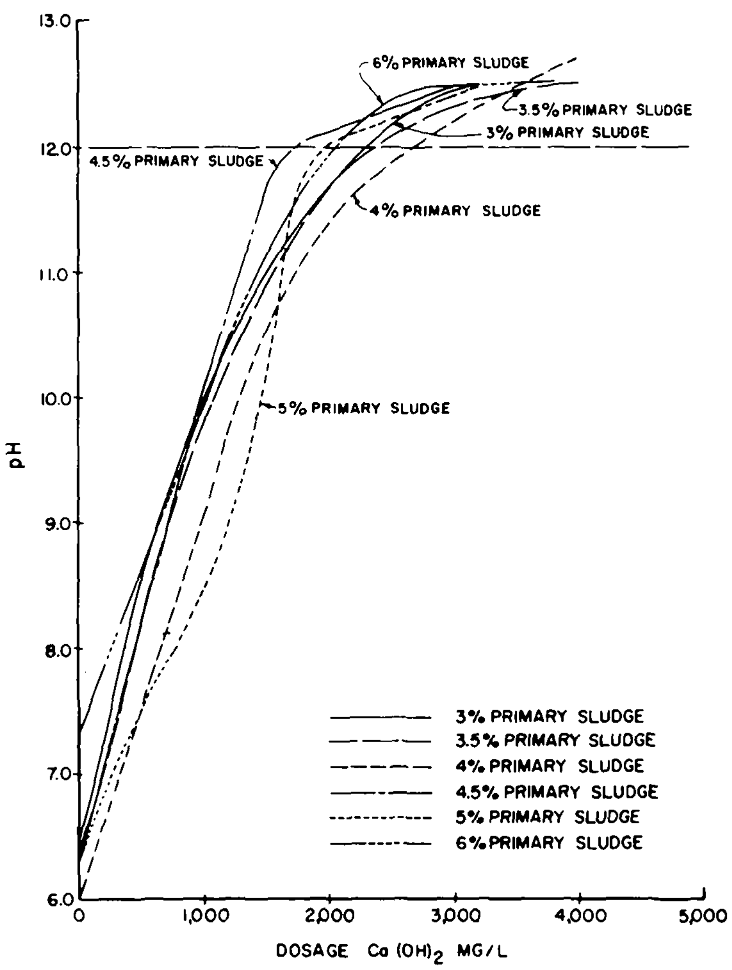



Figure 1-2.—Lime dosage versus pH primary sludge.

necticut. Wastewater treatment plant capacities which were representative ranged from 1 to 30 Mgal/d (0.04 to 1.31 $m^3/s$). A summary of these results has previously been reported.[10]

## Lime Requirements

The lime dosage required to exceed pH 12 for at least 30 minutes was found to be affected by the type of sludge, its chemical composition, and percent solids. As an operational procedure, a target pH of 12.5 was selected to insure that the final pH would be greater than 12. A summary of the lime dosage required for various sludges is shown in table 1–1. Of the total amount of lime which was required, an excess of 0 to 50 percent was added after pH 12 was reached in order to maintain the pH. Figure 1–1 shows the combined lime dosage versus pH for primary, anaerobically digested, waste activated, and septage sludges. Figures 1–2 to 1–5 describe the actual lime dosages which were required for each sludge type.

Table 1–2 compares the Lebanon full-scale test results, which are described later in the case history, with the data previously presented by Farrell et al., Counts, et al., and Paulsrud and Eikum for raw primary sludges. In general, excellent correlation was achieved.

Counts[9] has proposed the following equation for predicting the lime dosage required for primary and secondary sludges from the Richland, Wash., trickling filter plant:

Lime dose = 4.2 + 1.6 (TS)

When: Lime dose is expressed in grams $Ca(OH)_2$ per liter of sludge TS is the total solids fraction in the sludge.

Table 1–3 compares the values predicted by the Counts equation to the Lebanon data for raw primary, waste activated, anaerobically digested, and septage sludges.

With increasing solids concentrations, the Counts equation results in lower than actual lime dosages.

## pH Versus Time

Previous research has attempted to determine the magnitude of pH decay versus time and to quantify the

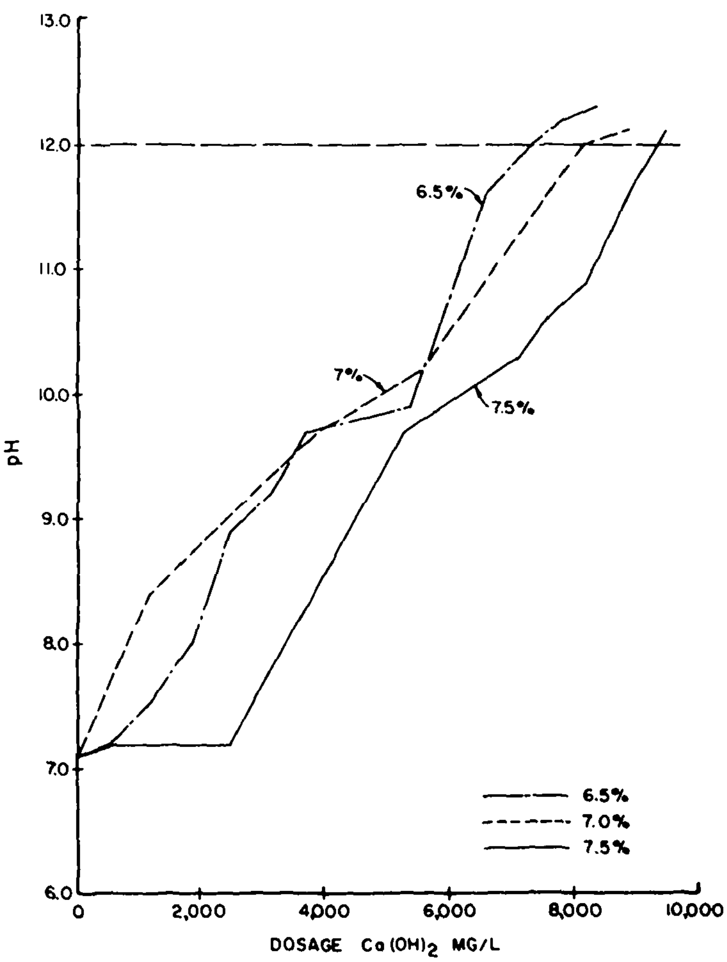



Figure 1–3.—Lime dosage versus pH anaerobic digested sludge.

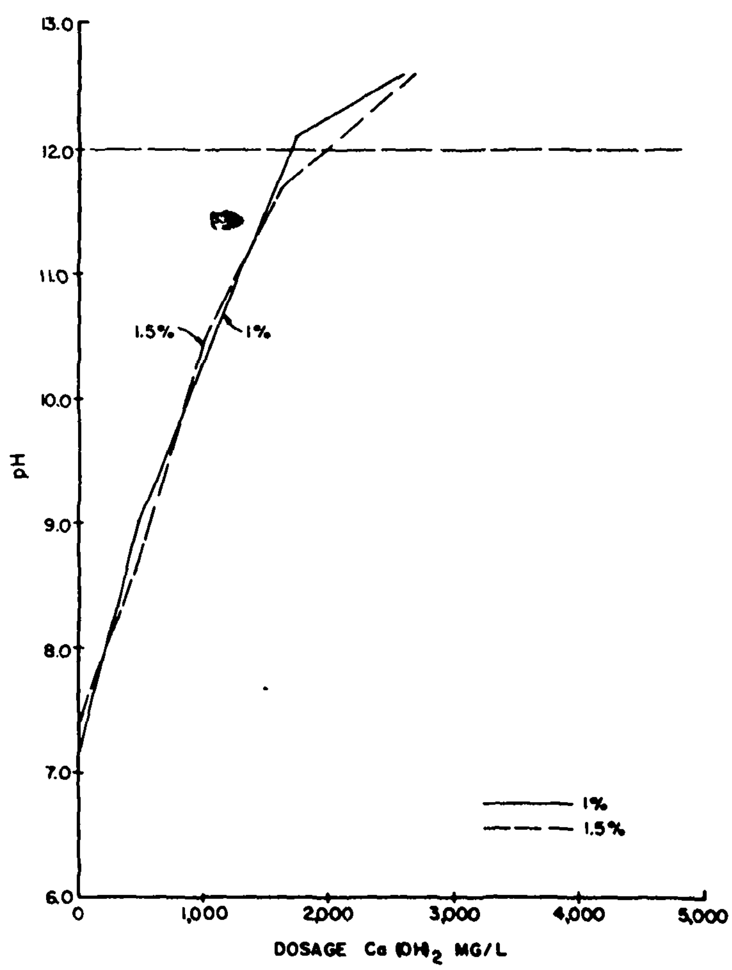



Figure 1–4.—Lime dosage versus pH waste activated sludge.

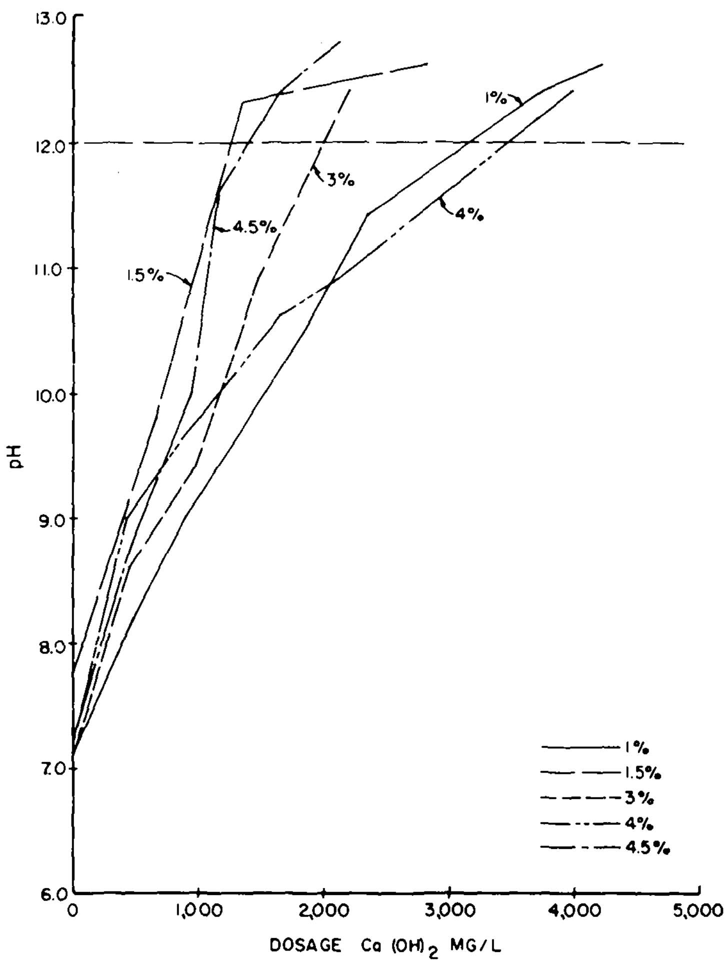



Figure 1–5.—Lime dosage versus pH septage.

Table 1–2.—Comparison of lime dosages required to treat raw primary sludge

| Investigator | Lime dose, kg lime/kg sludge dry solids |
|---|---|
| Burgess & Niple, Ltd (Lebanon) | [a]0.120 |
| Farrell et al | [b]0.098 |
| Counts et al | [c]0.086 |
| Paulsrud et al | [a]0.125 |

[a]Based on pH 12.5 for sludges reported.
[b]Based on pH 11.5 for sludges reported.
[c]Based on 4.78% solids.

variables which affect pH decay. Paulsrud[8] reported that negligible pH decay occurred when the sludge mixture was raised to pH 12 or greater or when the lime dose was approximately five times the dose to reach pH 11. In either case, for raw primary sludge, Paulsrud's dose was in the range of 0.100 to 0.150 kg lime/kg dry solids, which was approximately the dosage used at Lebanon.

Table 1–3.—Comparison of lime dosages predicted by the counts equation to actual data at Lebanon, Ohio

| Sludge type | Percent solids | Actual lime dose, kg lime/kg D.S. | Counts' lime dose, kg lime/kg D.S. |
|---|---|---|---|
| Raw primary | 4.78 | 0.120 | 0.086 |
| Waste activated | 1.37 | 0.300 | 0.305 |
| Anaerobically digested | 6.40 | 0.190 | 0.065 |
| Septage | 2.35 | 0.200 | 0.180 |

Counts[9] hypothesized that pH decay was caused by the sludge chemical demand which was exerted on the hydroxide ions supplied in the lime slurry. He further concluded that the degree of decay probably decreased as the treated sludge pH increased because of the extremely large quantities of lime required to elevate the pH to 12 or above. However, this pH phenomenon is probably because pH is an exponential function, e.g., the amount of $OH^-$ at pH 12 is ten times more than the amount of $OH^-$ at pH 11.

In the full-scale work at Lebanon, all sludges were lime stabilized to pH 12 or above and held for at least 30 minutes with the addition of excess lime. All treated sludges had less than a 2.0 pH unit drop after six hours. Limed primary sludge was the most stable with septage being the least stable. During the full-scale program, only the pH of limed primary sludge was measured for a period greater than 24 hours, which showed a gradual drop to approximately 11.6 after 18 hours beyond which no further decrease was observed.

The total mixing times from start through the 30 min contact time at Lebanon were as follows:

| | *Hours* |
|---|---|
| Primary sludge | 2.4 |
| Waste activated sludge | 1.7 |
| Septic tank sludge | 1.5 |
| Anaerobic digested sludge | 4.1 |

Mixing time was a function of lime slurry feed rate and was not limited by the agitating capacity of the diffused air system. Mixing time may have been reduced by increasing the capacity of the lime slurry tank.

To further examine the effects of excess lime addition above the levels necessary to reach pH 12, a series of laboratory tests were set up using a standard jar test apparatus. The tests were made on six one-liter portions of primary sludge with 2.7 percent total solids. The pH of each of the samples was increased to 12 by the addition of 10 percent hydrated lime slurry. One sample was used as a control. The remaining samples had 30 percent, 60 percent, 90 percent, 120 percent, and 150 percent by weight of the lime dose added to the control. The samples were mixed continuously for 6 hours and then again 10 minutes prior to each additional pH measurement. There was a negligible drop in pH over a

10-day period for those tests where excess lime was added.

A second laboratory scale test was completed using a 5 gal (19 l) raw primary sludge sample which was lime stabilized to pH 12.5 and allowed to stand at 18°C. Samples were withdrawn weekly and analyzed for pH and bacteria concentration. The results of the pH and bacteria studies are shown on figures 1–6 and 1–7, respectively. After 36 days, the pH had dropped to 12.0.

In conclusion, significant pH decay should not occur once sufficient lime has been added to raise the sludge pH to 12.5 and maintain that value for at least 30 min.

## Odors

Previous work[9] stated that the threshold odor number of raw primary and trickling filter sludges was approximately 8,000, while that of lime stabilized sludges usually ranged from 800 to 1,300. By retarding bacterial regrowth, the deodorizing effect can be prolonged. Further, it was concluded that by incorporating the stabilized sludge into the soil, odor potential should not be significant.

During the full-scale operations at Lebanon, there was an intense odor when raw sludge was first pumped to the lime stabilization mixing tank, which increased when diffused air was applied for mixing. As the sludge pH increased, the sludge odor was masked by the odor of ammonia which was being air stripped from the sludge. The ammonia odor was most intense with anaerobically digested sludge and was strong enough to cause nasal irritation. As mixing proceeded, the treated sludge acquired a musty humuslike odor, with the exception of septage which did not have a significant odor reduction as a result of treatment.

## Sludge Characteristics

Several authors have previously attempted to summarize the chemical and bacterial compositions of sewage sludges.[11–13] Recent data on the nutrient concentrations for various sludges have been reported by Sommers.[12] Chemical and pathogenic data on raw and lime stabilized raw primary, waste activated, septage, and anaero-

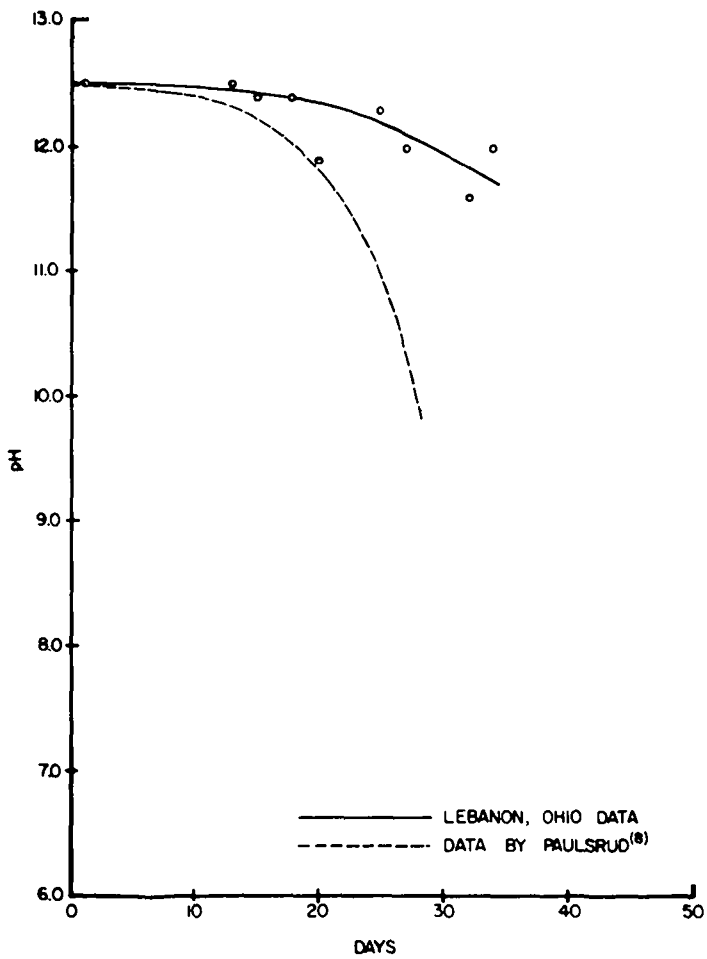



Figure 1–6.—Lime stabilized primary sludge pH versus time.

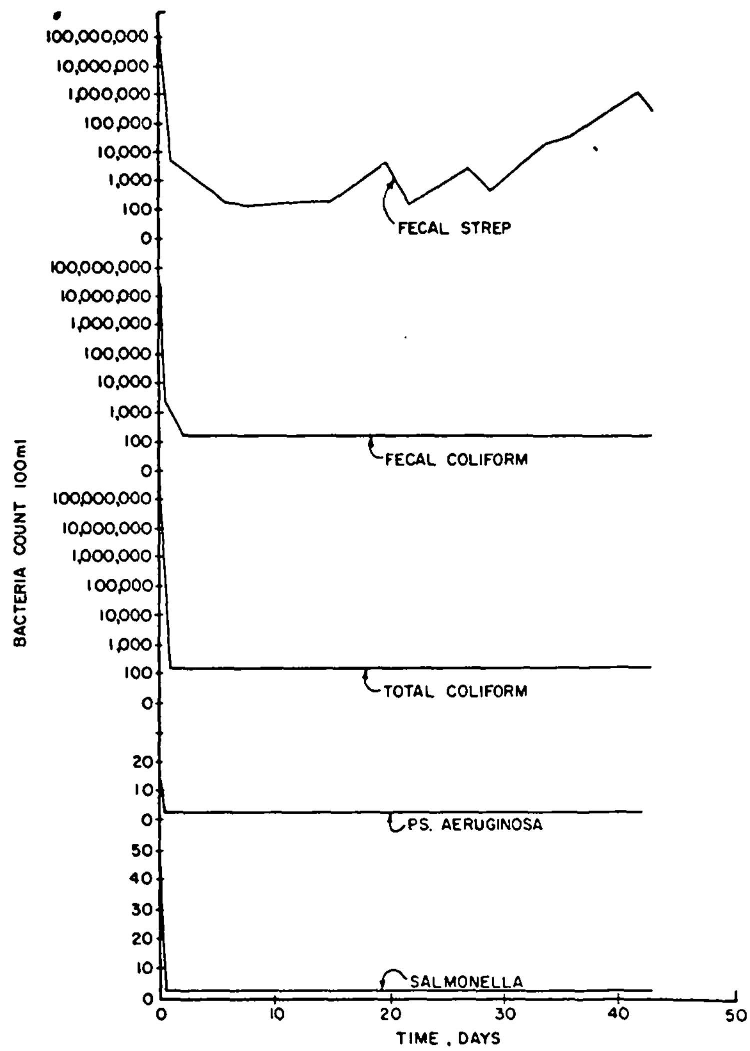



Figure 1–7.—Bacteria concentration versus time laboratory regrowth studies.

bically digested sludges from the Lebanon, Ohio full scale project have been summarized below and are included in more detail in the case history.

The addition of lime and mixing by diffused air altered the chemical characteristics of each sludge. In all sludges, lime stabilization resulted in an increase in alkalinity and soluble COD and a decrease in soluble phosphate. Total COD and total phosphate decreased for all sludges except waste activated. Ammonia nitrogen and total Kjeldahl nitrogen decreased for all sludges except waste activated.

The volatile solids concentrations of raw and lime stabilized sludges are shown in table 1–4. The actual volatile solids concentrations following lime stabilization are lower than those which would result only from the addition of lime. Neutralization, saponification, and hydrolysis reactions with the lime probably result in the lower volatile solids concentrations.

In terms of the agricultural value, lime stabilized sludges had lower soluble phosphate, ammonia nitrogen, total Kjeldahl nitrogen, and total solids concentrations than anaerobically digested primary/waste activated mixtures from the same plant, as shown in table 1–5. The significance of these changes is discussed in the section on land disposal.

Considerable research has been conducted on the degree of bacterial reduction which can be achieved by high lime doses.[14,15] In general, the degree of pathogen reduction increased as sludge pH increased with consistently high pathogen reductions occurring only after the pH reached 12.0. Fecal streptococci appeared to resist inactivation by lime treatment particularly well in the lower pH values; however, at pH 12, these organisms were also inactivated after 1 hour of contact time.[9]

In all lime stabilized sludges, *Salmonella* and *Pseudomonas aeruginosa* concentrations were reduced to near zero. Fecal and total coliform concentrations were reduced greater than 99.99 percent in the primary and septic sludges. In waste activated sludge, the total and fecal coliform concentrations decreased 99.9 percent and 99.94 percent, respectively. The fecal streptococci kills were as follows: primary sludge, 99.93 percent; waste activated sludge, 99.41 percent; septic sludge, 99.90 percent; and anaerobic digested, 96.81 percent.

Pathogen concentrations for the lime stabilized sludges are summarized in table 1–6.

Anaerobic digestion is currently an acceptable method of sludge stabilization.[16] For reference, lime stabilized sludge pathogen concentrations at Lebanon have been compared in table 1–6 to those observed for well digested sludge from the same plant.

Table 1–4.—Volatile solids concentration of raw and lime stabilized sludges

| Sludge type | Raw sludge volatile solids, solids concentration, mg/l | Lime stabilized sludge volatile solids, solids concentration, mg/l |
|---|---|---|
| Privary | 73.2 | 54.4 |
| Waste | 80.6 | 54.2 |
| Septage | 69.5 | 50.6 |
| Anaerobically digested | 49.6 | 37.5 |

Table 1–5.—Nitrogen and phosphorus concentrations in anaerobically digested and lime stabilized sludge

| Sludge type | Total phosphate as P, mg/l | Total Kjeldahl nitrogen as N, mg/l | Ammonia nitrogen as N, mg/l |
|---|---|---|---|
| Lime stabilized primary | 283 | 1,374 | 145 |
| Lime stabilized waste activated | 263 | 1,034 | 53 |
| Lime stabilized septage | 134 | 597 | 84 |
| Anaerobic digested | 580 | 2,731 | 709 |

Table 1–6.—Comparison of bacteria in anaerobic digested versus lime stabilized sludges

| | Fecal coliform #/100 ml | Fecal streptococci #/100 ml | Total coliform #/100 ml | Salmonella #/100 ml | Ps. aeruginosa #/100 ml |
|---|---|---|---|---|---|
| Anaerobically digested | $1,450 \times 10^3$ | $27 \times 10^3$ | $27,800 \times 10^3$ | 6 | 42 |
| Lime stabilized[a] | | | | | |
| Primary | $4 \times 10^3$ | $23 \times 10^3$ | $27.6 \times 10^3$ | [b]3 | 3 |
| Waste action | $16 \times 10^3$ | $61 \times 10^3$ | $212 \times 10^3$ | 3 | 13 |
| Septage | 265 | 665 | 2,100 | 3 | 3 |

[a]To pH equal to or greater than 12.0.
[b]Detection limit = 3.

Pathogen concentrations in lime stabilized sludges range from 10 to 1,000 times less than for anaerobically digested sludge.

A pilot scale experiment was completed in the laboratory to determine the viability and regrowth potential of bacteria in lime stabilized primary sludge over an extended period of time.

The test was intended to simulate storing stabilized sludge in a holding tank or lagoon when weather conditions prohibit spreading. In the laboratory test, 5 gal (19 l) of 7 percent raw sludge from the Mill Creek sewage treatment plant in Cincinnati were lime stabilized to pH 12.0. Lime was added until equivalent to 30 percent of the weight of the dry solids which resulted in a final pH of 12.5. The sample was then covered with foil and kept at room temperature 65°F (18.3°C) for the remainder of the test. The contents were stirred before samples were taken for bacterial analysis.

The results, shown on figure 1–7, indicate that a holding period actually increases the bacteria kill. *Salmonella* in the raw sludge totaling 44 per 100 ml were reduced to the detection limit by lime stabilization. *Pseudomonas aeruginosa* totaling 11 per 100 ml in the raw sludge were reduced to the detection limit by lime stabilization. The initial fecal coliform count of $3.0 \times 10^7$ was reduced to $5 \times 10^3$ after lime stabilization, and after 24 hours was reduced to less than 300. The raw sludge contained $3.8 \times 10^8$ total coliforms, but 24 hours after lime stabilization the coliform total was less than 300. The fecal strep count in the raw sludge was $1.8 \times 10^8$ which decreased to $9.6 \times 10^4$ after lime stabilization. After 24 hours, the count was down to $7.0 \times 10^3$ and after 6 days reduced to less than 300. The count increased to $8 \times 10^5$ after 40 days.

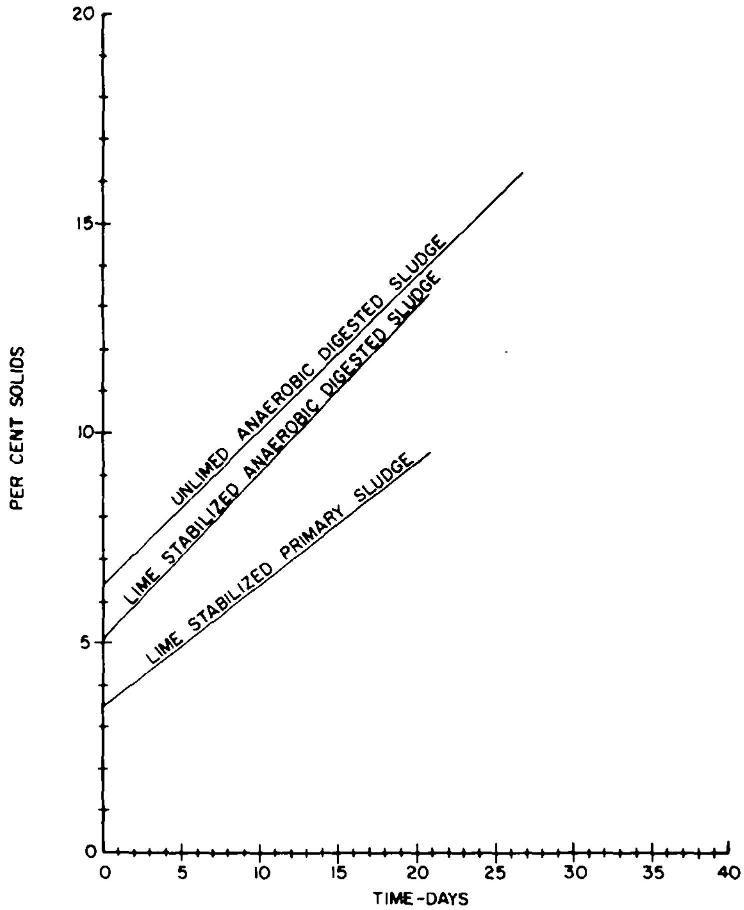



Figure 1–8.—Dewatering characteristics of various sludges on sand drying beds.

## Sludge Dewatering Characteristics

Farrell et al.[7] have previously reported on the dewatering characteristics of ferric chloride and alum treated sludges which were subsequently treated with lime. Trubnick and Mueller[17] presented, in detail, the procedures to be followed in conditioning sludge for filtration, using lime with and without ferric chloride. Sontheimer[18] presented information on the improvements in sludge filterability produced by lime addition.

Standard sand drying beds, which were located at the Lebanon, Ohio wastewater treatment plant, were used for sludge dewatering comparisons. Each bed was 30 ft × 70 ft (9.2 × 21.5 m). For the study, one bed was partitioned to form two, each 15 ft × 70 ft (4.6 × 21.5 m). Limed primary sludge was applied to one bed with limed anaerobically digested sludge being applied to the other side. A second full-sized bed was used to dewater unlimed anaerobically digested sludge. The results of the study are summarized on figure 1–8.

Lime stabilized sludges generally dewatered at a lower rate than well digested sludges. After 10 days, lime stabilized primary sludge had dewatered to approximately 6.5 percent solids as opposed to 9 percent for lime stabilized anaerobically digested sludge, and 10 percent for untreated anaerobically digested sludge.

The anaerobically digested sludge cracked first and dried more rapidly than either of the lime stabilized sludges. Initially, both of the lime stabilized sludges matted, with the digested sludge cracking after approximately 2 weeks. The lime stabilized primary sludge did not crack which hindered drying and resulted in the lower percent solids values.

## Land Application

Numerous references are available regarding the application of anaerobically digested sludges to agricultural land.[11,12,16,19] The application of sewage sludge on land has generally been viewed from two standpoints, either as a rate of application consistent with the utilization of nutrients in sludge by growing plants (i.e., agricultural utilization), or as the maximum amount of sludge applied in a minimum amount of time (i.e., disposal only). USEPA guidelines[16] generally favor the former approach. The successful operation of a program utilizing the application of sewage sludge on land is dependent upon a knowledge of the particular sludge, soil, and crop characteristics.

Organic matter content, fertilizer nutrients, and trace element concentrations are generally regarded as being vital to the evaluation of the applicability of land application of sewage sludge. The range of nitrogen, phosphorus, and potassium concentrations for sewage sludges have been reported by Brown et al.[11]

Sommers[12] has also summarized fertilizer recommendations for crops based primarily on the amount of major nutrients (nitrogen, phosphorus, and potassium) required by a crop and on the yield desired.

Counts[9] conducted greenhouse and test plot studies for lime stabilized sludges which were designed to provide information on the response of plants grown in sludge-soil mixtures ranging in application rate from 5 to 100 tons/acre (11 to 220 Mg/ha). Counts concluded that sludge addition to poor, e.g., sandy, soils would increase productivity, and therefore would be beneficial. The total nitrogen and phosphorous levels in plants grown in greenhouse pots, which contained sludge-soil mixtures, were consistently lower than plants which were grown in control pots. The control set, which contained only soil with no sludge additions, received optimum additions of chemical fertilizer during the actual plant growth phase of the studies. Calcium concentration in plant tissues from the sludge-soil pots were higher than those for the controls. The pH values of the various sludge-soil mixtures were lower after plant growth than before. Counts attributed the decrease to carbon dioxide buildup in the soil which resulted from biological activity.

Land application studies at Lebanon, Ohio, were conducted by spreading liquid sludge on agricultural land and on controlled test plots. Winter wheat, soybeans, and hay were grown on fields which were in normal agricultural production. Corn, swiss chard, and soybeans were grown on 22 test plots, each with an area of 0.021 acre (0.0085 ha). A preliminary report on the results of the land application studies will be published in 1978.[29]

Sludge application was accomplished by spreading as a liquid using a four-wheel drive vehicle which was equipped with a 600 gal (2.3 $m^3$) tank. The width of sludge spread per pass was approximately 24 in. (60 cm).

The lime stabilized sludge formed a filamentous mat 1/8–1/4 in (0.3–0.6 cm) thick which, when dry, partly choked out the wheat. The mat partly deteriorated over time, but significant portions remained at the time of harvest. There was no matting on the fields where the lime stabilized sludge was incorporated into the soil before planting.

Spontaneous growth of tomatoes was significant in the fields which had lime stabilized sludge incorporated into the soil before planting. Seeds were contained in the sludge and were not sterilized by the lime. These plants were absent at the site where the sludge was not incorporated, even though no herbicide was applied, probably because of frequent frosts and the lack of sludge incorporation into the soil. During the next year's growing season, an increase in insect concentration was noticed on the fields which had received lime stabilized sludge.

## LIME STABILIZATION DESIGN CONSIDERATIONS

### Overall Design Concepts

Lime and sludge are two of the most difficult materials to transfer, meter, and treat in any wastewater treatment plant. For these reasons, design of stabilization facilities should emphasize simplicity, straightforward piping layout, ample space for operation and maintenance of equipment, and gravity flow wherever possible. As discussed in more detail in the following sections, lime transport should be by auger with the slurry or slaking operations occurring at the point of use. Lime slurry pumping should be avoided with transport being by gravity in open channels. Sludge flow to the tank truck and/or temporary holding lagoon should also be by gravity if possible.

Figures 1–9, 1–10, and 1–11 show conceptual designs for lime stabilization facilities at wastewater treatment facilities with 1, 5, and 10 Mgal/d (0.04, 0.22 and 0.44 $m^3/s$) throughputs. The 1 Mgal/d (0.04 $m^3/s$) plant, as shown on figure 1–9, utilizes hydrated lime and a simple batch mixing tank, with capability to treat all sludges in less than one shift per day. Treated sludge could be allowed to settle for several hours before hauling in order to thicken, and thereby reduce the volume hauled. Alternately, the sludge holding lagoon could be used for thickening.

Figure 1–10 shows the conceptual design for lime stabilization facilities of a 5 Mgal/d (0.22 $m^3/s$) wastewater treatment facility. Pebble lime is utilized in this installation. Two sludge mixing tanks are provided, each with the capacity to treat the total sludge production from two shifts. During the remaining shift, sludge could be thickened and hauled to the land disposal site. Alternately, the temporary sludge lagoon could be used for sludge thickening.

Figure 1–11 shows the conceptual design for lime stabilization facilities of a 10 Mgal/d (0.44 $m^3/s$) wastewater treatment plant. A continuous lime treatment tank with 2 hours detention time is used to raise the sludge pH to 12. A separate sludge thickening tank is provided to increase the treated sludge solids content before land application. Sludge transport is assumed to be by pipeline to the land disposal site. A temporary sludge holding lagoon was assumed to be necessary, and would also be located at the land disposal site.

### Lime Requirements

The quantity of lime which will be required to raise the pH of municipal wastewater sludges to pH greater than 12 can be estimated from the data presented in table 1–1 and from figures 1–2 to 1–5. Lime dosages have been shown as 100 percent $Ca(OH)_2$ and should be adjusted for the actual type of lime used. Generally, the lime requirements for primary and/or waste activated sludge will be in the range of 0.1 to 0.3 Kg 100 percent $Ca(OH)_2$ per Kg of dry sludge solids. Laboratory jar testing can confirm the dosage required for existing sludges.

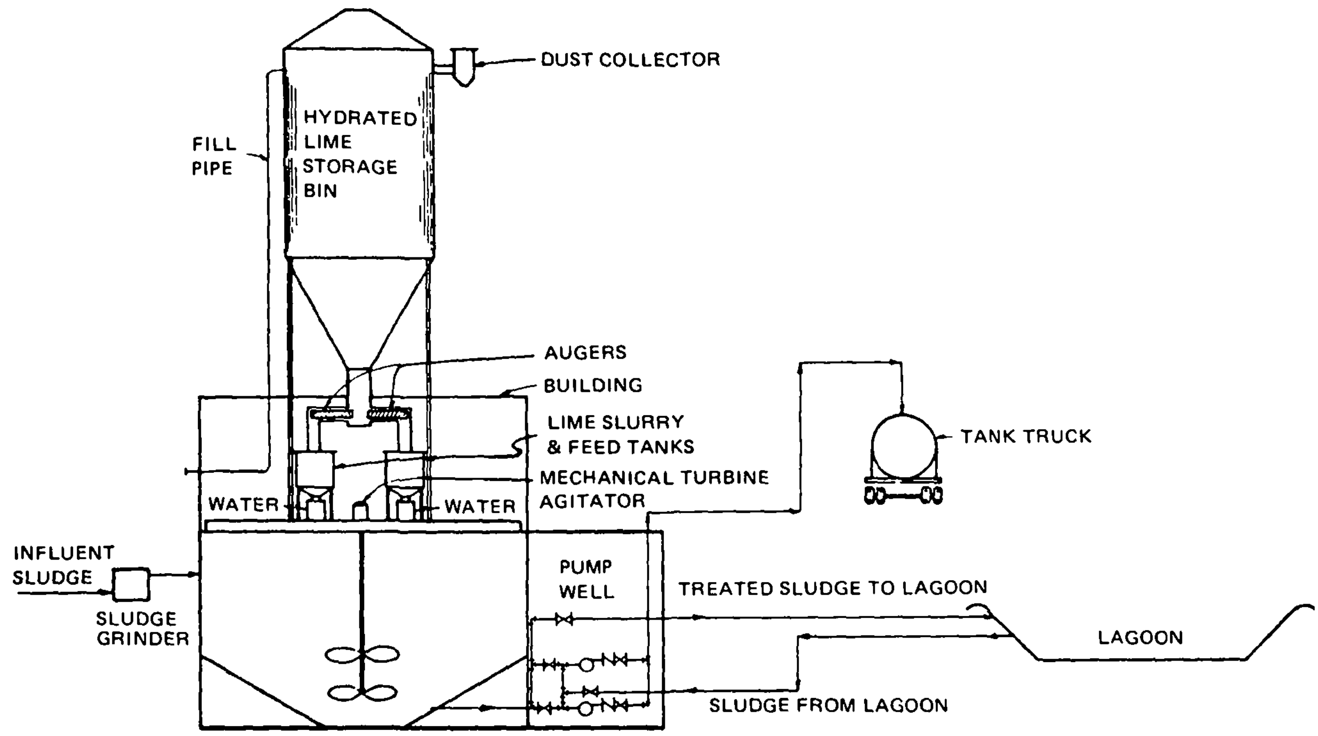



Figure 1–9.—Conceptual design for lime stabilization facilities for a 1 Mgal/d (0.04 $m^3$/s) treatment plant.

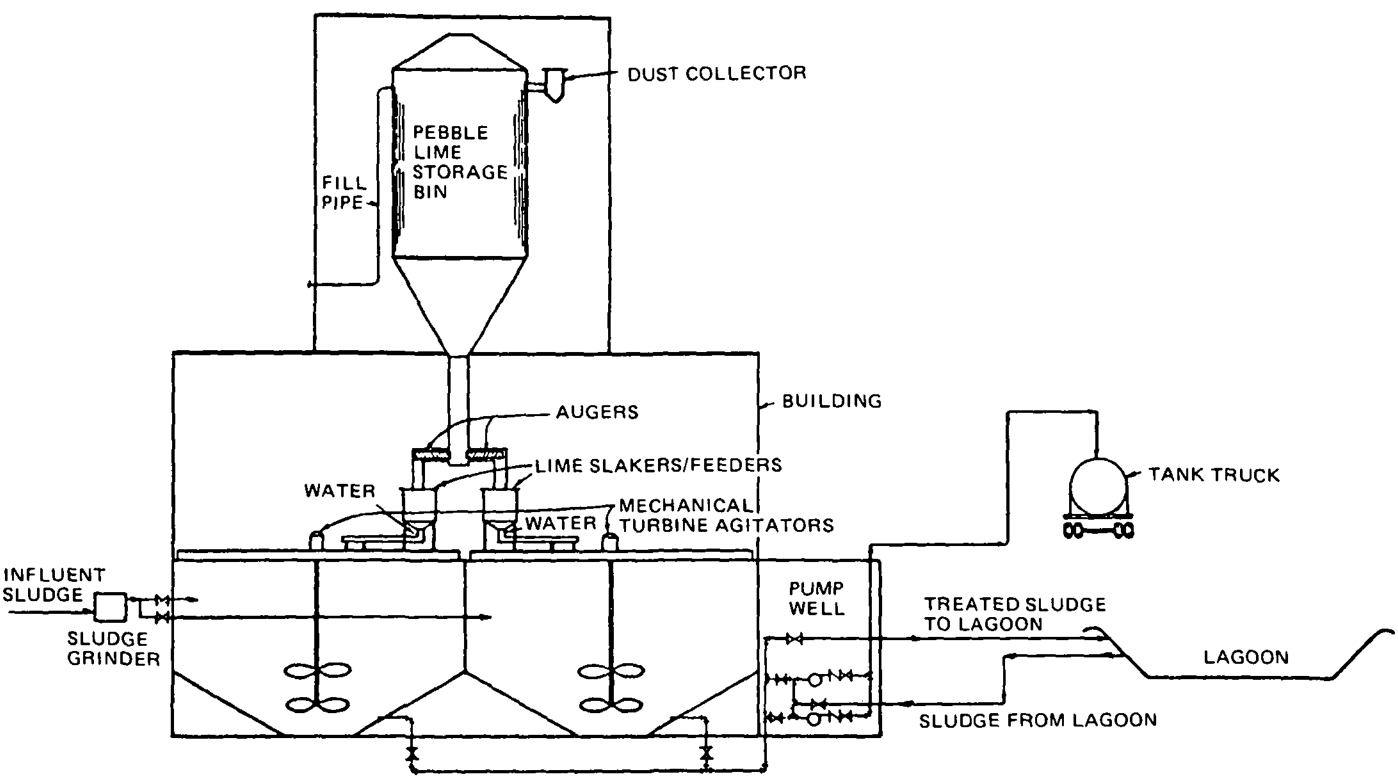



Figure 1–10.—Conceptual design for lime stabilization facilities for a 5 Mgal/d (0.22 $m^3$/s) treatment plant.

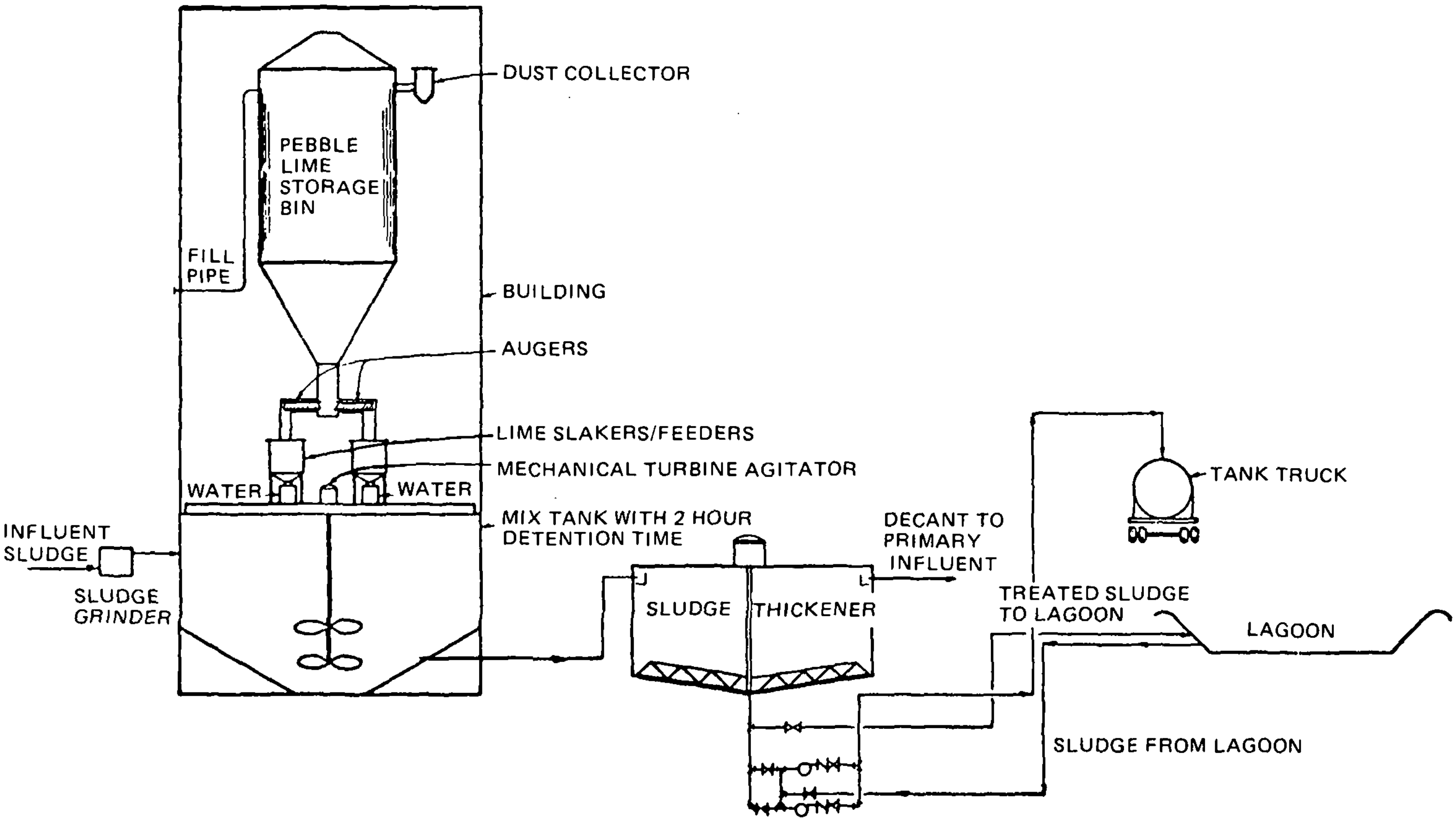



Figure 1-11.—Conceptual design for lime stabilization facilities for a 10 Mgal/d (0.44 $m^3/s$) treatment plant.

## Types of Lime Available

Lime in its various forms, as quicklime and hydrated lime, is the principal, lowest cost alkali. Lime is a general term, but by strict definition, it only embraces burned forms of lime—quicklime, hydrated lime, and hydraulic lime. The two forms of particular interest to lime stabilization, however, are quicklime and hydrated lime. Not included are carbonates (limestone or precipitated calcium carbonate) that are occasionally but erroneously referred to as "lime."[20]

### *Quicklime*

Quicklime is the product resulting from the calcination of limestone and to a lesser extent shell. It consists primarily of the oxides of calcium and magnesium. On the basis of their chemical analyses, quicklimes may be divided into three classes:

1. High calcium quicklime—containing less than 5 percent magnesium oxide, 85–90 percent CaO
2. Magnesium quicklime—containing 5 to 35 percent magnesium oxide, 60–90 percent CaO
3. Dolomitic quicklime—containing 35 to 40 percent magnesium oxide, 55–60 percent CaO

The magnesium quicklime is relatively rare in the United States and, while available in a few localities, is not generally obtainable.

Quicklime is available in a number of more or less standard sizes, as follows:

1. Lump lime—the product with a maximum size of 8 in (20 cm) in diameter down to 2 in (5.1 cm) to 3 in (7.6 cm) produced in vertical kilns.
2. Crushed or pebble lime—the most common form, which ranges in size from about 2-1/4 in (5.1–0.6 cm), produced in most kiln types.
3. Granular lime—the product obtained from Fluo-Solids kilns that has a particulate size range of 100 percent passing a #8 sieve and 100 percent retained on a #80 sieve (a dustless product).
4. Ground lime—the product resulting from grinding the larger sized material and/or passing off the fine size. A typical size is substantially all passing a #8 sieve and 40 to 60 percent passing a #100 sieve.
5. Pulverized lime—the product resulting from a more intense grinding that is used to produce ground lime. A typical size is substantially all passing a #20 sieve and 85 to 95 percent passing a #100 sieve.

6. Pelletized lime—the product made by compressing quicklime fines into about 1-inch size pellets or briquettes.

### *Hydrated Lime*

As defined by the American Society for Testing and Materials, hydrated lime is: "A dry powder obtained by treating quicklime with water enough to satisfy its chemical affinity for water under the conditions of its hydration."

The chemical composition of hydrated lime generally reflects the composition of the quicklime from which it is derived. A high calcium quicklime will produce a high calcium hydrated lime obtaining 72 percent to 74 percent calcium oxide and 23 percent to 24 percent water in chemical combination with the calcium oxide. A dolomitic quicklime will produce a dolomitic hydrate. Under normal hydrating conditions, the calcium oxide fraction of the dolomitic quicklime completely hydrates, but generally only a small portion of the magnesium oxide hydrates (about 5 to 20 percent). The composition of a normal dolomitic hydrate will be 46 percent to 48 percent calcium oxide, 33 percent to 34 percent magnesium oxide, and 15 percent to 17 percent water in chemical combination with the calcium oxide. (With some soft-burned dolomitic quicklimes, 20 percent to 50 percent of the MgO will hydrate.)

A "special" or pressure hydrated dolomitic lime is also available. This lime has almost all (more than 92 percent) of the magnesium oxide hydrated; hence, its water content is higher and its oxide content lower than the normal dolomitic hydrate.

Hydrated lime is packed in paper bags weighing 50 lb (22.7 kg) net; however, it is also shipped in bulk.

Quicklime is obtainable in either bulk carloads or tanker trucks or in 80 lb (36.4 kg) multiwall paper bags. Lump, crushed, pebble, or pelletized lime, because of the large particle sizes, is rarely handled in bags and is almost universally shipped in bulk. The finer sizes of quicklime, ground, granular, and pulverized, are readily handled in either bulk or bags.

## Lime Storage and Feeding

Depending on the type of lime, storage and feeding can be either in bag or bulk. Bagged lime will probably be more economical for treatment plants less than one Mgal/d (0.04 $m^3$/s) and for temporary or emergency feed systems, e.g., when a digester is out of service for cleaning and repair. In new facilities, bulk storage will probably be cost effective. Storage facilities should be constructed such that dry lime is conveyed to the point of use and then mixed or slaked. Generally, augers are best for transporting either hydrated or pebble lime. Auger runs should be horizontal or not exceeding an incline of 30°.[27]

The feeder facilities, i.e., dry feeder and slaking or slurry tank, should be located adjacent to the stabilization mixing tank such that lime slurry can flow by gravity in open channel troughs to the point of mixing. Pumping lime slurry should be avoided. Slurry transfer distances should be kept to a minimum. Access to feeder, slaker and/or slurry equipment should be adequate for easy disassembly and maintenance.

## Mixing

Lime/sludge mixtures can be mixed either with mechanical mixers or with diffused air. The level of agitation should be great enough to keep sludge solids suspended and dispense the lime slurry evenly and rapidly. The principal difference between the resultant lime stabilized sludges in both cases is that ammonia will be stripped from the sludge with diffused air mixing. Mechanical mixing has been used by previous researchers for lime stabilization but only on the pilot scale.

With diffused air mixing, adequate ventilation should be provided to dissipate odors generated during mixing and stabilization. Coarse bubble diffusers should be used with air supplies in the range of 150–250 $ft^3$/min per 1,000 $ft^3$ (150–250 $m^3$/min per 1,000 $m^3$) of mixing tank volume. Diffusers should be mounted such that a spiral roll is established in the mixing tank away from the point of lime slurry application. Diffusers should be accessible and piping should be kept against the tank wall to minimize the collection of rags, etc. Adequate piping support should be provided.

With the design of mechanical mixers, the bulk velocity (defined as the turbine agitator pumping capacity divided by the cross sectional area of the mixing vessel) should be in the range of 15 to 26 ft/min (4.6 to 7.9 m/min). Impeller Reynolds numbers should exceed 1,000 in order to achieve a constant power number.[21] The mixer should be specified according to the standard motor horsepower and AGMA gear ratios in order to be commercially available.

For convenience, table 1–7 was completed which shows a series of tank and mixer combinations which should be adequate for mixing sludges up to 10 percent dry solids, over a range of viscosity, and Reynolds number combinations which were as follows:

Max. Reynolds number 10,000 at 100 cp sludge viscosity
Max. Reynolds number 1,000 at 1,000 cp sludge viscosity

Table 1–7 can be used to select a mixer horsepower and standard AGMA gear combination depending on the volume of sludge to be stabilized. For example, for a 5,000 gal (1.9 $m^3$) tank, any of the mixer-turbine combinations should provide adequate mixing. Increasing turbine diameter and decreasing shaft speed results in a decrease in horespower requirement as shown.

Additional assumptions were that the bulk fluid velocity must exceed 26 ft/min (7.9 m/min), impeller Reynolds number must exceed 1,000, and that power requirements

Table 1-7.—Mixer specifications for sludge slurries

| Tank size, liters | Tank diameter, meters | Prime mover, hp/shaft speed, r/min | Turbine diameter, centimeters |
|---|---|---|---|
| 18,925 | 2.9 | 7.5/125 | 81 |
| | | 5/84 | 97 |
| | | 3/56 | 109 |
| 56,775 | 4.2 | 20/100 | 114 |
| | | 15/68 | 135 |
| | | 10/45 | 160 |
| | | 7.5/37 | 170 |
| 113,550 | 5.3 | 40/84 | 145 |
| | | 30/68 | 155 |
| | | 25/56 | 168 |
| | | 20/37 | 206 |
| 283,875 | 7.2 | 100/100 | 157 |
| | | 75/68 | 188 |
| | | 60/56 | 201 |
| | | 50/45 | 221 |
| 378,500 | 8.0 | 125/84 | 183 |
| | | 100/68 | 198 |
| | | 75/45 | 239 |

ranging from 0.5–1.5 hp/1,000 gal (0.5–1.5 hp per 3,785 l) are necessary. The mixing tank configuration assumed that the liquid depth equals tank diameter and that baffles with a width of 1/12 the tank diameter were placed at 90° spacing. Mixing theory and equations which were used were after Badger,[21] Hicks,[22] and Fair.[23]

## Raw and Treated Sludge Piping, Pumps, and Grinder

Sludge piping design should include allowances for increased friction losses due to the non-Newtonian properties of sludge. Friction loss calculations should be based on treated sludge solids concentrations and should allow for thickening in the mixing tank after stabilization. Pipelines should not be less than 2 in (5.1 cm) in diameter and should have tees in major runs at each change in direction to permit rodding, cleaning, and flushing the lines. Adequate drains should be provided. If a source of high pressure water is available (either nonpotable or noncross-connected potable), it can be used to flush and clean lines.

Spare pumps should be provided and mounted such that they can be disassembled easily. Pump impeller type materials of construction should be adequate for the sludge solids concentration and pH.

Sludge grinding equipment should be used to make the raw sludge homogenous. Sticks, rags, plastic, etc., will be broken up prior to lime stabilization to improve the sludge mixing and flow characteristics and to eliminate unsightly conditions at the land disposal site.

# A CASE HISTORY OF LIME STABILIZATION

## Background

Facilities for lime stabilization of sludge were incorporated into an existing 1.0 Mgal/d (0.04 $m^3$/s) single stage activated sludge wastewater treatment plant located at Lebanon, Ohio. Lebanon has a population of about 8,000 and is located in southwestern Ohio, 30 mi (48 km) northeast of Cincinnati. The surrounding area is gently rolling farmland with a small number of light industries, nurseries, orchards, and truck farms.

Major unit processes at the wastewater treatment plant include influent pumping, preaeration, primary clarification, conventional activated sludge, and anaerobic sludge digestion. Average influent $BOD_5$ and suspended solids concentrations are 180 and 243 mg/l, respectively. The treatment plant flow schematic is shown on figure 1-12.

Prior to completing the sludge liming system, the existing anaerobic sludge digester was inoperative and was being used as a sludge holding tank. The digester pH was approximately 5.5 to 6.0. Grit and sand accumulations had reduced its effective volume to 40–50 percent of the total. Waste activated sludge was being returned to the primary clarifiers and resettled with the primary sludge. Combined primary/waste activated sludge was being pumped to the digester and ultimately recycled to the primary clarifiers via the digester supernatant. Typical supernatant suspended solids concentrations were in the range of 30,000 to 40,000 mg/l. When possible, sludge was withdrawn from the digester and dewatered on sand drying beds.

USEPA made the decision to utilize lime stabilization at Lebanon not only as a full scale research and demonstration project, but also as a means of solids handling during the period while the anaerobic digester was out of service for cleaning and repair.

## Revisions to the Existing Wastewater Treatment Plant

### *Lime Stabilization*

The lime stabilization process was designed to treat raw primary, waste activated, septic tank, and anaerobically digested sludges. The liming system was integrated with the existing treatment plant facilities, as shown on figure 1-13. Hydrated lime was stored in a bulk storage bin and was augered into a volumetric feeder. The feeder transferred dry lime at a constant rate into a 25 gal (95 l) slurry tank which discharged an 8–10 percent lime slurry by gravity into an existing 6,500 gal (25 $m^3$) tank. The lime slurry and sludge were mixed with diffused air. A flow schematic for the lime stabilization facilities is shown on figure 1-14. Design data are shown in table 1-8.

### *Septage Holding Facilities*

Because the Lebanon wastewater treatment plant routinely accepted septic tank pumpings, a 5,000 gal (18.4

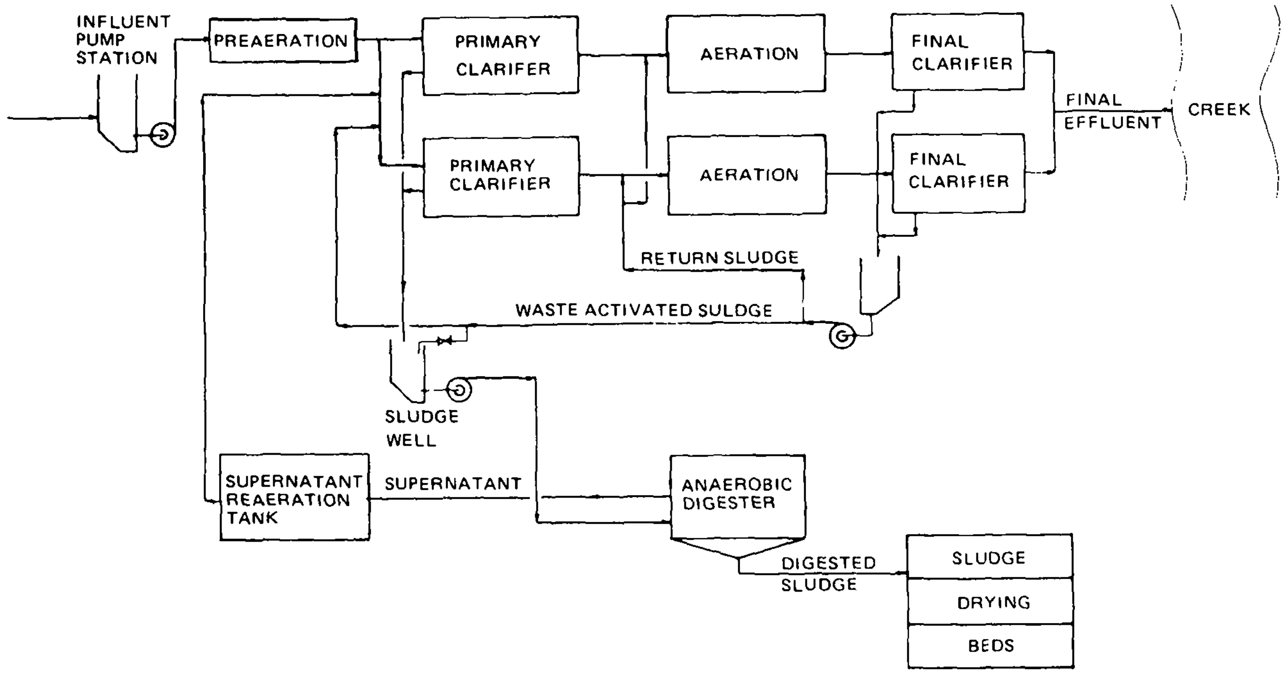



Figure 1–12.—Treatment plant flow schematic prior to incorporating lime stabilization.

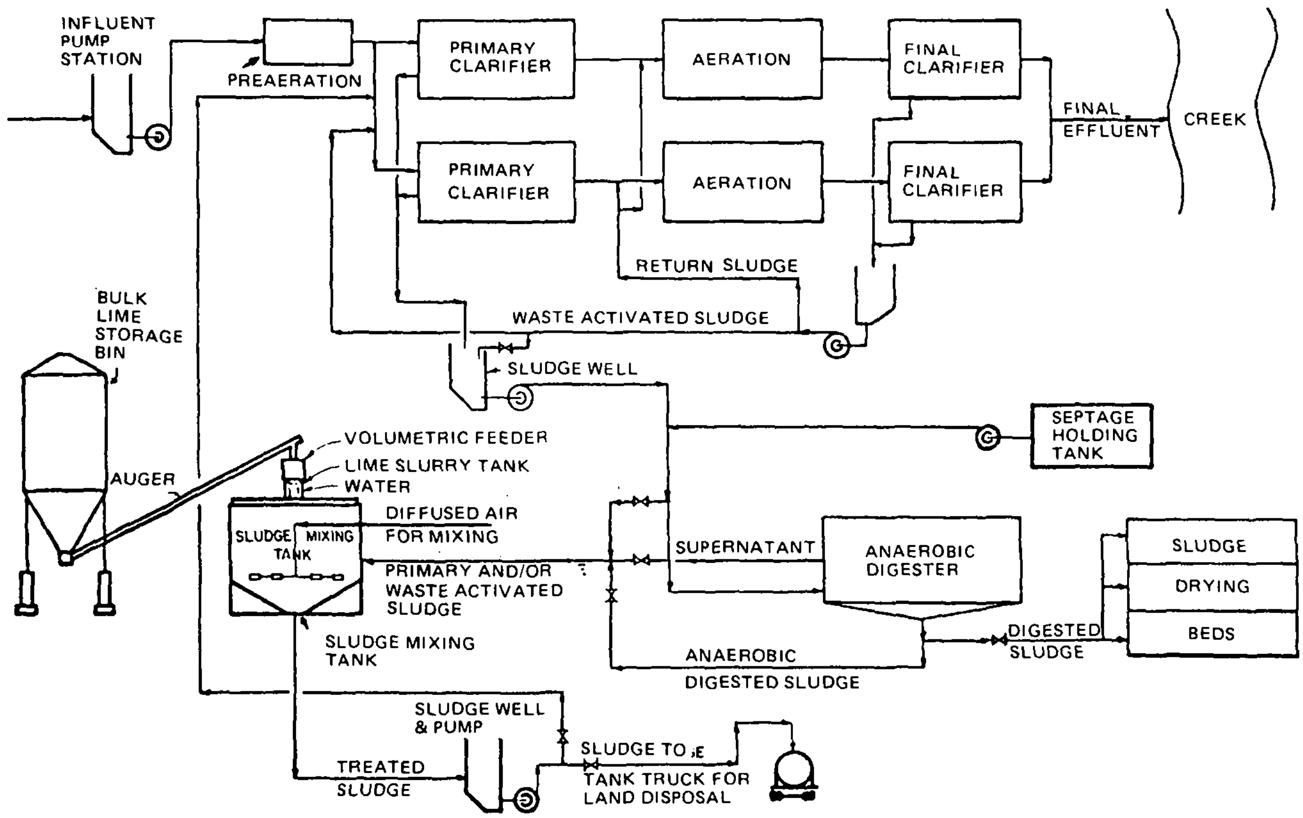



Figure 1–13.—Treatment plant flow schematic after incorporating lime stabilization.

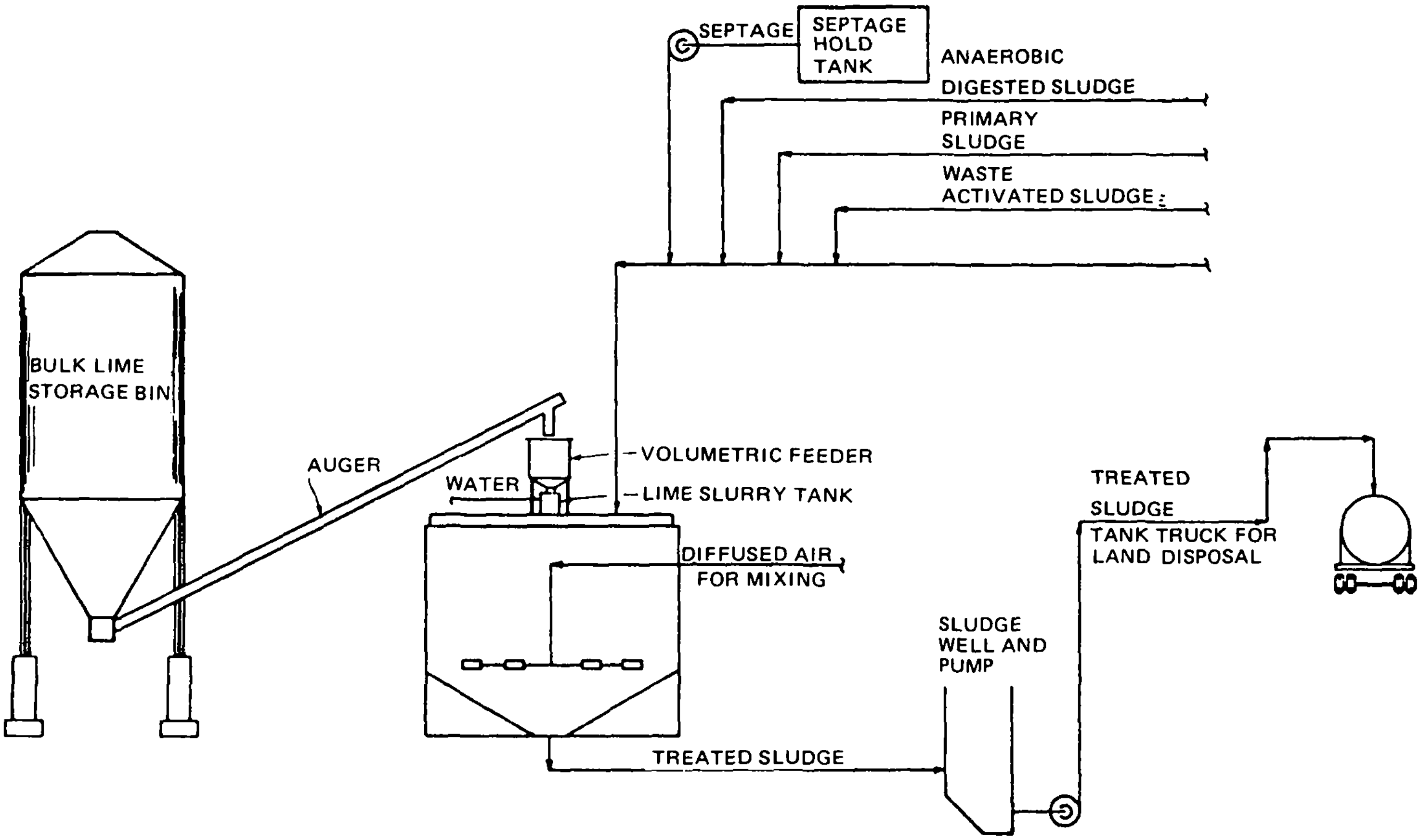



Figure 1–14.—*Lime stabilization process flow diagram.*

$m^3$) tank was installed to hold septic tank sludges prior to lime treatment. The tank was equipped with a transfer pump which could be used to either feed the lime stabilization process or transfer septage to the primary tank influent at a controlled rate.

## *Ultimate Sludge Disposal*

Treated sludges were applied to sand drying beds, to test plots, and to three productive agricultural sites. Land spreading operations began in early March and continued through October. The sludge hauling vehicle was a four-wheel drive truck with a 600 gal (2.3 $m^3$) tank.

## *Anaerobic Digester*

As previously described, the existing single stage anaerobic sludge digester was inoperative and was being used as a sludge holding tank. The digester and auxiliary equipment were completely renovated and returned to good operating condition which allowed a comparison of anaerobic digestion and lime stabilization. The digester was cleaned, a new boiler and hot water circulating system were installed, and all necessary repairs were made to piping, valves, pumps, and electrical equipment.

The anaerobic digester design data are shown in table 1–9.

## Operation and Sampling

Raw sludge, e.g., primary, waste activated, septage or digested sludge, was pumped to the mixing tank where it was mixed by diffused air. Four coarse bubble diffusers were mounted approximately 1 ft (30.5 cm) above the top of the tank hopper and 1.25 ft (38 cm) from the tank wall. This location permitted mixing to roll sludge up and across the tank at which point lime slurry was fed. Lime which was used for the stabilization of all sludges was industrial grade hydrated lime with CaO and MgO contents of 46.9 percent and 34 percent, respectively. All lime requirements have been converted and are expressed as 100 percent $Ca(OH)_2$ except as noted. Samples were taken from the untreated, but thoroughly mixed, sludge for chemical, pH, bacteria, and parasite analyses.

After the initial pH determination, the lime slurry addition was started. Hydrated lime was augered from the lime storage bin to the volumetric feeder which was located directly above the sludge mixing tank. The lime was slurried by the tangential injection of water into a

Table 1–8.—Design data for lime stabilization facilities

| | |
|---|---|
| Mixing tank | |
| Total volume | 30 $m^3$ (8,000 gal) |
| Working volume | 25 $m^3$ (6,500 gal) |
| Dimensions | 3.05 m × 3.66 m × 2.38 m (10′ × 12′ × 7.8′) |
| Hoppered bottom | 0.91 m (3′) @ 27° slope |
| Type of diffuser | Coarse bubble |
| Number of diffusers | 4 |
| Air supply | 14–34 $m^3$/min (500–1,200 $ft^3$/min) |
| Bulk lime storage | |
| Total volume | 28 $m^3$ (1,000 $ft^3$) |
| Diameter | 2.74 m (9′) |
| Vibrators | 2 ea Syntron V-41 |
| Fill system | Pneumatic |
| Discharge system | 15 cm (6″) dia. auger |
| Material of construction | Steel |
| Type and manufacturer | Columbian model C-95 |
| Volumetric feeder | |
| Total volume | 0.28 $m^3$ (10 cu ft) |
| Diameter | 71 cm (28″) |
| Material of construction | Steel |
| Type and manufacturer | Vibrascrew LBB 28–10 |
| Feed range | 45–227 kg/hr (100–500 lb/hr) |
| Average feed rate | 78 kg/hr (173 lb/hr) |
| Lime slurry tank | |
| Total volume | 94.6 l (25 gal) |
| Diameter | 0.61 m (2′) |
| Septic tank sludge holding tank (septage tank) | |
| Total volume | 18.4 $m^3$ (650 $ft^3$) |
| Working volume | 15 $m^3$ (4,000 gal) |
| Dimensions | 3.66 m × 1.92 m × 2.62 m (12′ × 6.3′ × 8.6′) |
| Mixing | Coarse bubble |
| Number of diffusers | 1 |
| Air supply | 2.8–8.4 $m^3$/min (100–300 $ft^3$/min) |
| Transfer pumps | |
| Raw and treated sludge | 1,136 l/min (300 gpm) |
| Septage transfer pump | 379 l/min (100 gpm) |

Table 1–9.—Anaerobic digester rehabilitation design data

| | |
|---|---|
| Tank dimensions | 15 m (50′) dia. × 6.1 m (20′) SWD |
| Total volume | 1,223 $m^3$ (43,200 $ft^3$) |
| Actual volatile solids loading | 486 g VSS/$m^3$ (0.03 lb VSS/$ft^3$) |
| Hydraulic detention time | 36 days |
| Sludge recirculation rate | 757 1/min (200 gpm) |
| Boiler capacity | 2.53 × $10^8$ Joules/hr (240,000 Btu/hr) |

25 gal (95 l) slurry tank. The lime solution (8–10 percent by weight) then flowed by gravity into an open channel with three feed points into the sludge mixing tank.

The sludge pH was checked every 15 minutes as the lime slurry was added until the sludge reached a pH of 12, at which time it was held for 30 minutes. During the 30 minute period, lime slurry continued to be added. After 30 minutes, samples were taken for chemical, bacteria, and parasite analyses. Air mixing was then discontinued, allowing the limed sludge to concentrate. The sludge then flowed by gravity to a sludge well from which it was pumped to the land disposal truck.

Samples of raw and treated Lebanon sludges were taken during each operating day of the lime stabilization operations. Anaerobically digested sludge samples were taken at the same time and analyzed for use in comparisons of chemical, bacterial, and pathogen properties.

Sample preservation and chemical analysis techniques were performed in accordance with procedures as stated in "Methods for Chemical Analysis of Water and Wastes, USEPA,"[24] and "Standard Methods for the Examination of Water and Wastewater."[25]

*Salmonella* species and *Pseudomonas aeruginosa* were determined by EPA staff using the method developed by Kenner and Clark.[26] Fecal coliform, total coliform, and fecal streptoccocus were determined according to methods specified in Standard Methods for Examination of Water and Wastewater.

## Raw Sludges

Chemical data for Lebanon, Ohio raw primary, waste activated, anaerobically digested, and septage sludges have been summarized in table 1–10. Data for each parameter include the average and range of the values observed.

Analyses for heavy metals were conducted on grab samples of raw primary, waste activated, and anaerobically digested sludges. These data have been reported in table 1–11 as mg/kg on a dry weight basis and include the average and range of values.

Pathogen data for Lebanon, Ohio raw primary, waste activated, anaerobically digested, and septage sludges have been summarized in table 1–12. In general, the data are in agreement with the values reported by Stern,[14] with the exception of *Salmonella* and *Pseudomonas aeruginosa*, which are lower than the reported values.

## Lime Stabilized Sludges

Chemical and bacterial data for lime stabilized sludges have previously been summarized in the general discussion on lime stabilization. Specific data from the Lebanon, Ohio full scale project have been summarized in tables 1–13 and 1–14. Lime stabilized sludges had lower soluble phosphate, ammonia nitrogen, total Kjeldahl nitrogen, and total solids concentrations than anaerobically digested primary/waste activated mixtures from the same plant.

In all lime stabilized sludges, *Salmonella* and *Pseudomonas aeruginosa* concentrations were reduced to near zero. Fecal and total coliform concentrations were reduced greater than 99.99 percent in the primary and septic sludges. In waste activated sludge, the total and fecal coliform concentrations decreased 99.99 percent

Table 1–10.—Chemical composition of raw sludges at Lebanon, Ohio

| Parameter (mg/l) | Raw primary sludge | Waste activated sludge |
|---|---|---|
| Alkalinity | 1,885 | 1,265 |
| Alkalinity range | 1,264–2,820 | 1,220–1,310 |
| Total COD | 54,146 | 12,810 |
| Total COD range | 36,930–75,210 | 7,120–19,270 |
| Soluble COD | 3,046 | 1,043 |
| Soluble COD range | 2,410–4,090 | 272–2,430 |
| Total phosphate, as P | 350 | 218 |
| Total phosphate range, as P | 264–496 | 178–259 |
| Soluble phosphate, as P | 69 | 85 |
| Soluble phosphate range, as P | 20–150 | 40–119 |
| Total Kjeldahl nitrogen | 1,656 | 711 |
| Total Kjeldahl nitrogen range | 1,250–2,470 | 624–860 |
| Ammonia nitrogen | 223 | 51 |
| Ammonia nitrogen range | 19–592 | 27–85 |
| Total suspended solids | 48,700 | 12,350 |
| Total suspended solids range | 37,520–65,140 | 9,800–13,860 |
| Volatile suspended solids | 36,100 | 10,000 |
| Volatile suspended solids range | 28,780–43,810 | 7,550–12,040 |
| Volatile acids | 1,997 | NA |
| Volatile acids range | 1,368–2,856 | NA |
| Alkalinity | 3,593 | 1,897 |
| Alkalinity range | 1,330–5,000 | 1,200–2,690 |
| Total COD | 66,372 | 24,940 |
| Total COD range | 39,280–190,980 | 10,770–32,480 |
| Soluble COD | 1,011 | 1,223 |
| Soluble COD range | 215–4,460 | 1,090–1,400 |
| Total phosphate, as P | 580 | 172 |
| Total phosphate range, as P | 379–862 | 123–217 |
| Soluble phosphate, as P | 15 | 25 |
| Soluble phosphate range, as P | 6.9–34.8 | 21.6–27.9 |
| Total Kjeldahl nitrogen | 2,731 | 820 |
| Total Kjeldahl nitrogen range | 1,530–4,510 | 610–1,060 |
| Ammonia nitrogen | 709 | 92 |
| Ammonia nitrogen range | 368–1,250 | 68–116 |
| Total suspended solids | 61,140 | 21,120 |
| Total suspended solids range | 48,200–68,720 | 6,850–44,000 |
| Volatile suspended solids | 33,316 | 12,600 |
| Volatile suspended solids range | 27,000–41,000 | 3,050–30,350 |
| Volatile acids | 137 | 652 |
| Volatile acids range | 24–248 | 560–888 |

and 99.47 percent, respectively. The fecal streptococci kills were as follows: primary sludge, 99.93 percent; waste activated sludge, 99.41 percent; septic sludge, 99.90 percent; and anaerobic digested sludge, 96.81 percent. Pathogen concentrations in lime stabilized sludges range from 10 to 1,000 times less than for anaerobically digested sludges.

## Economic Analysis

### *Lebanon Facilities*

As previously described, the anaerobic sludge digestion facilities at Lebanon were essentially inoperable at

Table 1–11.—Heavy metal concentrations in raw sludges at Lebanon, Ohio

| Metal (mg/kg) | Raw primary sludge | Waste activated sludge | Anaerobic digested sludge |
|---|---|---|---|
| Cadmium, average | 105 | 388 | 137 |
| Cadmium, range | 69–141 | 119–657 | 73–200 |
| Total chromium, average | 633 | 592 | 882 |
| Total chromium, range | 287–979 | 133–1,050 | 184–1,580 |
| Copper, average | 2,640 | 1,340 | 4,690 |
| Copper, range | 2,590–2,690 | 670–2,010 | 4,330–5,050 |
| Lead, average | 1,379 | 1,624 | 1,597 |
| Lead, range | 987–1,770 | 398–2,850 | 994–2,200 |
| Mercury, average | 6 | 46 | 0.5 |
| Mercury, range | 0.4–11 | 0.1–91 | 0.1–0.9 |
| Nickel, average | 549 | 2,109 | 388 |
| Nickel, range | 371–727 | 537–3,680 | 263–540 |
| Zinc, average | 4,690 | 2,221 | 7,125 |
| Zinc, range | 4,370–5,010 | 1,250–3,191 | 6,910–7,340 |

Table 1–12.—Pathogen data for raw sludges at Lebanon, Ohio

| Parameter (#/100 ml) | Raw primary sludge | Waste activated sludge |
|---|---|---|
| Salmonella average | 62 | 6 |
| Salmonella range | 11–240 | 3–9 |
| Ps. aeruginosa average | 195 | $5.5 \times 10^3$ |
| Ps. aeruginosa range | 75–440 | $91–1.1 \times 10^4$ |
| Fecal coliform average MF | NA | $2.65 \times 10^7$ |
| Fecal coliform range MF | NA | $2.0 \times 10^7–3.3 \times 10^7$ |
| Fecal coliform average MPN | $8.3 \times 10^8$ | NA |
| Fecal coliform range MPN | $1.3 \times 10^8–3.3 \times 10^9$ | NA |
| Total coliform average MF | NA | $8.33 \times 10^8$ |
| Total coliform range MF | NA | $1.66 \times 10^8–1.5 \times 10^9$ |
| Total coliform average MPN | $2.9 \times 10^9$ | NA |
| Total coliform range MPN | $1.3 \times 10^9–3.5 \times 10^9$ | NA |
| Fecal streptococci average | $3.9 \times 10^7$ | $1.03 \times 10^7$ |
| Fecal streptococci range | $2.6 \times 10^7–5.2 \times 10^7$ | $5 \times 10^5–2 \times 10^7$ |
| **Parameter (#/100 ml)** | **Anaerobically digested sludge** | **Septage sludge** |
| Salmonella average | 6 | 6 |
| Salmonella range | 3–30 | 3–9 |
| Ps. aeruginosa average | 42 | 754 |
| Ps. aeruginosa range | 3–240 | $14–2.1 \times 10^7$ |
| Fecal coliform average MF | $2.6 \times 10^5$ | $1.5 \times 10^7$ |
| Fecal coliform range MF | $3.4 \times 10^4–6.6 \times 10^5$ | $1.0 \times 10^7–1.8 \times 10^7$ |
| Fecal coliform average MPN | $1.45 \times 10^6$ | NA |
| Fecal coliform range MPN | $1.9 \times 10^5–4.9 \times 10^6$ | NA |
| Total coliform average MF | $2.42 \times 10^7$ | $2.89 \times 10^8$ |
| Total coliform range MF | $1.3 \times 10^5–1.8 \times 10^8$ | $1.8 \times 10^7–7 \times 10^8$ |
| Total coliform average MPN | $2.78 \times 10^7$ | NA |
| Total coliform range MPN | | NA |
| Fecal streptococci average | $2.7 \times 10^5$ | $6.7 \times 10^5$ |
| Fecal streptococci range | | $3.3 \times 10^5–1.2 \times 10^6$ |

Table 1–13.—Chemical composition of lime stabilized sludges at Lebanon, Ohio

| Parameter (mg/l) | Raw primary sludge | Waste activated sludge | Anaerobically digested sludge | Septage sludge |
|---|---|---|---|---|
| Alkalinity | 4,313 | 5,000 | 8,467 | 3,475 |
| Alkalinity range | 3,830–5,470 | 4,400–5,600 | 2,600–13,200 | 1,910–6,700 |
| Total COD | 41,180 | 14,700 | 58,690 | 17,520 |
| Total COD range | 26,480–60,250 | 10,880–20,800 | 27,190–107,060 | 5,660–23,900 |
| Soluble COD | 3,556 | 1,618 | 1,809 | 1,537 |
| Soluble COD range | 876–6,080 | 485–3,010 | 807–2,660 | 1,000–1,970 |
| Total phosphate | 283 | 263 | 381 | 134 |
| Total phosphate range | 164–644 | 238–289 | 280–460 | 80–177 |
| Soluble phosphate | 36 | 25 | 2.9 | 2.4 |
| Soluble phosphate range | 17–119 | 17–31 | 1.4–5.0 | 1.4–4.0 |
| Total Kjeldahl nitrogen | 1,374 | 1,034 | 1,980 | 597 |
| Total Kjeldahl nitrogen range | 470–2,510 | 832–1,430 | 1,480–2,360 | 370–760 |
| Ammonia nitrogen | 145 | 64 | 494 | 110 |
| Ammonia nitrogen range | 81–548 | 36–107 | 412–570 | 53–162 |
| Total suspended solids | 38,370 | 10,700 | 66,350 | 23,190 |
| Total suspended solids range | 29,460–44,750 | 10,745–15,550 | 46,570–77,900 | 14,250–29,600 |
| Volatile suspended solids | 23,480 | 7,136 | 26,375 | 11,390 |
| Volatile suspended solids range | 19,420–26,450 | 6,364–8,300 | 21,500–29,300 | 5,780–19,500 |

Table 1–14.—Pathogen data for lime stabilized sludges at Lebanon, Ohio

| Parameter (#/100 ml) | Raw primary sludge | Waste activated sludge | Anaerobically digested sludge | Septage sludge |
|---|---|---|---|---|
| Salmonella average | [a]3 | [a]3 | [a]3 | [a]3 |
| Salmonella range | [a]3 | [a]3 | [a]3 | [a]3 |
| Ps. aeruginosa average | [a]3 | [a]3 | [a]3 | [a]3 |
| Ps. aeruginosa range | [a]3 | [a]3–26 | [a]3 | [a]3 |
| Fecal coliform MF average | NA | $1.62 \times 10^4$ | $3.3 \times 10^3$ | $2.65 \times 10^2$ |
| Fecal coliform MF range | NA | $3.3 \times 10^2$–$3.2 \times 10^4$ | $3.3 \times 10^3$ | $2 \times 10^2$–$3.3 \times 10^2$ |
| Fecal coliform average MPN | $5.93 \times 10^3$ | NA | 18 | NA |
| Fecal coliform range MPN | 560–$1.7 \times 10^4$ | NA | 18 | NA |
| Total coliform average MF | NA | $2.2 \times 10^5$ | NA | $2.1 \times 10^3$ |
| Total coliform range MF | NA | $3.3 \times 10^3$–$4.2 \times 10^5$ | NA | 200–$4 \times 10^3$ |
| Total coliform average MPN | $1.15 \times 10^5$ | NA | 18 | NA |
| Total coliform range MPN | 640–$5.4 \times 10^5$ | NA | 18 | NA |
| Fecal streptococci average | $1.62 \times 10^4$ | $6.75 \times 10^3$ | $8.6 \times 10^3$ | 665 |
| Fecal streptococci range | $4.0 \times 10^3$–$5.5 \times 10^4$ | $1.5 \times 10^3$–$1.35 \times 10^3$ | $3.3 \times 10^3$–$1.4 \times 10^4$ | $3.3 \times 10^2$–$1 \times 10^3$ |

[a]Detectable limit = 3.

the start of the lime stabilization project. Funds were allocated to construct lime stabilization facilities, as well as to rehabilitate the anaerobic digester. In both cases, the existing structures, equipment, etc., were utilized to the maximum extent possible. Table 1–15 includes the actual amounts paid to contractors, following competitive bidding, and does not include engineering fees, administrative costs, etc.

The cost of the lime stabilization facilities was $29,507.45 compared to $32,134.59 for cleaning and repair of the anaerobic sludge digester.

## Capital Cost of New Facilities

Capital and annual operation and maintenance costs for lime stabilization and anaerobic sludge digestion facilities were estimated assuming new construction as a part of a 1.0 Mgal/d (0.04 $m^3/s$) wastewater treatment plant with primary clarification and single stage conventional activated sludge treatment processes.

The capital costs for lime stabilization facilities included a bulk lime storage bin for hydrated lime, auger, volumetric feeder and lime slurry tank, sludge mixing and

Table 1–15.—Actual cost of digester rehabilitation and lime stabilization facilities construction

| | |
|---|---|
| Anaerobic digester cleaning | |
| Cleaning contractor | $5,512.12 |
| Temporary sludge lagoon | 2,315.20 |
| Lime for stabilizing digester contents | 514.65 |
| Temporary pump rental | 300.30 |
| Subtotal digester cleaning | 8,642.27 |
| Anaerobic digester rehabilitation | |
| Electrical equipment, conduit, etc | 1,055.56 |
| Natural gas piping | 968.76 |
| Hot water boiler, piping, pump, heat exchanger repair | 7,472.26 |
| Control room rehabilitation | 1,465.00 |
| Sludge recirculating pump repair | 771.00 |
| Piping and valve rehabilitation | 8,587.30 |
| Floating cover roof repair | 1,014.04 |
| Repair utilities, drains | 211.52 |
| Miscellaneous | 1,946.88 |
| Subtotal digester rehabilitation | 23,492.32 |
| Lime stabilization process | |
| Electrical equipment, conduit, etc | 1,692.00 |
| 3″ and 4″ sludge lines, supports, valves, and fittings | 6,140.19 |
| 4″ sludge crossover pipe, valves, and fittings | 1,101.48 |
| 1-1/2″ air line and diffusers | 1,310.00 |
| 3/4″ water lines and hose bibbs | 865.00 |
| Lime bin, auger, vibrators | 7,229.44 |
| Volumetric feeder, trough and gate | 3,460.00 |
| Existing pump repairs | 3,399.00 |
| Miscellaneous metal | 1,200.00 |
| Relocate sanitary service line | 200.00 |
| Repair utilities | 134.00 |
| Miscellaneous | 934.34 |
| Contractor's overhead | 1,842.00 |
| Subtotal lime stabilization | 29,507.45 |
| Septage holding tank | |
| Septage holding tank and pump | 6,174.70 |
| Subtotal septage holding tank | 6,174.70 |
| Total cost for digester cleaning and rehabilitation, lime stabilization, and septage facilities | 67,816.74 |

thickening tank with a mechanical mixer, sludge grinder, all weather treatment building, electrical and instrumentation, interconnecting piping and transfer pumps, and 60-day detention treated sludge holding lagoon. The basis for design is as follows:

| | |
|---|---|
| Daily primary sludge dry solids production | 1,250 lbs/day (568 kg/day) |
| Average primary sludge volume @ 5 percent solids | 2,910 gal/day (11 $m^3$/day) |
| Daily waste activated dry solids production | 1,084 lbs/day (493 kg/day) |
| Average waste activated sludge volume @ 1.5 percent solids | 8,580 gal/day (32 $m^3$/day) |
| Average lime dosage required per unit dry solids | 0.20 kg/kg |
| Daily lime requirement as 100 percent $Ca(OH)_2$ | 475 lb/day (215 kg/day) |
| Treatment period | 3 hrs/day |
| Bulk lime storage bin volume minimum | 1,000 $ft^3$ (28 $m^3$) |
| Bulk lime storage bin detention time | 34 days |
| Lime feeder and slurry tank capacity (spared) | 5–15 $ft^3$/hr (0.14–0.42 $m^3$/hr) |
| Influent sludge grinder capacity | 200 gpm (12.6 l/s) |
| Sludge mixing tank volume | 15,000 gal (57 $m^3$) |
| Sludge mixing tank dimensions | 14 ft × 14 ft × 10 ft SWD (4.3 m × 4.3 m × 3 m) |
| Sludge mixer horsepower | 15 HP (11.2 kW) |
| Sludge mixer turbine diameter | 53 in (135 cm) |
| Turbine speed | 68 rpm |
| Sludge transfer pump capacity (spared) | 400 gpm (25.2 l/s) |
| Treated sludge percent solids | 4 percent |
| Sludge holding lagoon volume | 100,000 $ft^3$ (2,860 $m^3$) |
| Sludge holding lagoon maximum detention time | 60 days |
| Treatment building floor area | 150 $ft^2$ (14 $m^2$) |
| Treatment building construction | Brick and block |
| Instrumentation | pH record treated sludge volume |

Capital costs for the lime stabilization facilities were based on July 1, 1977, bid date, and have been summarized in table 1–16.

Lime stabilization operation assumed one man, 2 hours per day, 365 days per year, at $6.50 per hour, including overhead. Maintenance labor and materials assumed 52 hours per year labor at $6.50 per hour and $800 per year for maintenance materials. The total quantity of 46.8 percent CaO hydrated lime required was 141 tons (128 Mg) per year at $44.50 per ton ($49.06/Mg)

The total annual cost for lime stabilization, excluding land application of treated sludge, has been summarized in table 1–17.

The basis for design of a single stage anaerobic sludge digester for the same treatment plant was as follows:

| | |
|---|---|
| Daily primary sludge dry solids production | 1,250 lb/day (568 kg/day) |

Table 1–16.—Capital cost of lime stabilization facilities for a new 1 Mgal/d (0.04 $m^3$/s) wastewater treatment plant

| | |
|---|---|
| Site work, earthwork and yard piping | $6,000 |
| Lime storage bin and feeders | 30,000 |
| Treatment tank, pumps, sludge grinders, and building structure | 52,000 |
| Electrical and instrumentation | 10,000 |
| Sludge holding lagoon | 20,000 |
| Subtotal construction cost | 118,000 |
| Engineering | 12,000 |
| Total capital cost | 130,000 |
| Amortized cost at 30 yrs., 7% int. (CRF = 0.081) | 10,500 |
| Annual capital cost per ton dry solids | 24.65 |

Table 1–17.—Total annual cost for lime stabilization excluding land disposal for a 1 Mgal/d (0.04 $m^3$/s) plant

| Item | Total annual cost | Annual cost per Kkg dry solids | Annual cost per ton dry solids |
|---|---|---|---|
| Operating labor | $4,700 | $12.14 | $11.03 |
| Maintenance labor and materials | 1,100 | 2.84 | 2.58 |
| Lime | 6,300 | 16.20 | 14.74 |
| Laboratory | 500 | 1.29 | 1.17 |
| Capital | 10,500 | 27.11 | 24.65 |
| Total annual cost | 23,100 | 59.58 | 54.17 |

| | |
|---|---|
| Average primary sludge volume @ 5 percent solids | 2,910 gal/day (11 $m^3$/day) |
| Daily waste activated dry solids production | 1,084 lb/day (493 kg/day) |
| Average waste activated sludge volume @ 1.5 percent solids | 8,580 gal/day (32 $m^3$/day) |
| Daily volatile solids production | 1,634 lb/day (743 kg/day) |
| Volatile solids loading | 0.05 lb VSS/$ft^3$/day (0.8 kg/$m^3$/day) |
| Digester hydraulic detention time | 21 days |
| Digester gas production | 13 $ft^3$/lb VSS (0.8 $m^3$/kg) feed |
| Average volatile solids reduction | 50 percent |
| Digested sludge dry solids production | 1,515 lb/day (690 kg/day) |
| Digested sludge percent solids | 6 percent |
| Digester net heat requirement | 186,000 Btu/hr (54,500 W) |
| Mechanical mixer horsepower | 15 HP (11.2 kW) |
| Sludge recirculation pumps (2 ea) | 350 gal/min ea (22 l/s) |

Capital cost for the anaerobic sludge digestion facilities, including the control building, structure, floating cover, heat exchanger, gas safety equipment, pumps, and interconnecting piping, assuming July 1, 1977, bid date, and engineering, legal, and administrative costs are summarized in table 1–18.

Digester operation assumed one man, 1 hour per day, 365 days per year at $6.50 per hour, including overhead. Maintenance labor and material assumed 52 hours per year at $6.50 per hour and $1,500 per year for maintenance materials.

The cost of anaerobic digester operation was offset by assuming a value of $2.10 per million Btu ($1.99 per million kJ) for all digester gas produced above the net digester heat requirement.

The total annual cost for anaerobic sludge digestion, excluding land application has been summarized in table 1–19.

Both the lime stabilization and anaerobic digestion alternatives were assumed to utilize land application of treated sludge as a liquid hauled by truck. The capital cost for a sludge hauling vehicle was assumed to be $35,000, which was depreciated on a straight line basis over a 10-year period. Alternatively, a small treatment plant could utilize an existing vehicle which could be converted for land application at a somewhat lower capital cost.

The assumed hauling distance was 3 to 5 miles (5 to 8 km), round trip. Hauling time assumed 10 minutes to fill, 15 minutes to empty, and 10 minutes driving, or a total of 35 minutes per round trip. The truck volume was assumed to be 1,500 gal (5.68 $m^3$) per load. The cost of truck operations, excluding the driver and depreciation, was assumed to be $8.50 per operating hour. The truck driver labor rate was assumed to be $6.50 per hour, including overhead.

Truck operation time was based on hauling an average of 6,860 gal (1.812 $m^3$) of lime stabilized sludge, i.e., five loads and 2,940 gal (0.777 $m^3$) of anaerobically digested sludge, i.e., two loads per day. The reduced volume of anaerobically digested sludge resulted from the volatile solids reduction during digestion and the higher solids concentration compared to lime stabilized sludge.

Although it may be possible to obtain the use of farmland at no cost, e.g., on a voluntary basis, the land application economic analysis assumed that land would

Table 1–18.—Capital cost for single stage anaerobic digestion facilities for a 1 Mgal/min (0.04 $m^3$/s) wastewater treatment plant

| | |
|---|---|
| Site work, earthwork, yard piping | $44,000 |
| Digester | 233,000 |
| Control building | 133,000 |
| Electrical and instrumentation | 47,000 |
| Subtotal construction cost | 457,000 |
| Engineering | 46,000 |
| Total capital cost | 503,000 |
| Amortized cost at 30 yrs, 7% int. (CRF = 0.081) | 40,700 |
| Annual capital cost per unit feed dry solids | 95.54 |

Table 1–19.—Total annual cost for single stage anaerobic sludge digestion excluding land disposal for a 3,785 $m^3$/day plant

| Item | Total annual cost | Annual cost per Kkg dry solids | Annual cost per ton dry solids |
|---|---|---|---|
| Operating labor | $2,400 | $6.20 | $5.63 |
| Maintenance labor and materials | 1,800 | 4.65 | 4.23 |
| Laboratory | 500 | 1.29 | 1.17 |
| Capital | 40,700 | 105.09 | 95.54 |
| Fuel credit | (2,900) | (7.49) | (6.81) |
| Total annual cost | 42,500 | 109.74 | 99.76 |

Table 1–20.—Annual cost for land application of lime stabilized and anaerobically digested sludges for a 3,785 m$^3$/day plant

| Item | Lime stabilization | | | Anaerobic digestion | | |
|---|---|---|---|---|---|---|
| | Total annual cost | Annual cost per Kkg solids | Annual cost per ton solids | Total annual cost | Annual cost per Kkg solids | Annual cost per ton solids |
| Amortized cost of land | $2,600 | $6.75 | $6.14 | $1,700 | $4.39 | $3.99 |
| Truck depreciation | 3,500 | 9.04 | 8.22 | 3,500 | 9.04 | 8.22 |
| Truck driver | 7,100 | 18.35 | 16.67 | 2,800 | 7.24 | 6.57 |
| Truck operation | 9,300 | 24.03 | 21.83 | 3,600 | 9.30 | 8.45 |
| Laboratory | 500 | 1.29 | 1.17 | 500 | 1.29 | 1.17 |
| Fertilizer credit | (3,100) | (8.05) | (7.30) | (2,000) | (8.05) | (7.30) |
| Land credit | (2,200) | (5.68) | (5.16) | (1,400) | (3.62) | (3.29) |
| Total annual cost | 17,700 | 45.73 | 41.57 | 8,700 | 19.59 | 17.81 |

be purchased at a cost of $750 per acre ($1850 per ha). Sludge application rates were assumed to be 10 dry tons per acre per year (22.4 Mg/ha/year). Land costs were amortized at 7 percent interest over a 30-year period.

To offset the land cost, a fertilizer credit of $7.30 per ton ($8.05 per Mg) of dry sludge solids was assumed. This rate was arbitrarily assumed to be 50 percent of the value published by Brown[11] based on medium fertilizer market value and low fertilizer content. The reduction was made to reflect resistance to accepting sludge as fertilizer. The land cost was further offset by assuming a return of $50 per acre ($124 ha), either as profit after farming expenses, or as the rental value of the land.

Capital and annual operation and maintenance costs for land application of lime stabilized and anaerobically digested sludges have been summarized in table 1–20.

For each item in table 1–18, the total annual cost was calculated and divided by the total raw primary plus waste activated sludge quantity, i.e., 426 tons/year (386 Mg). Anaerobically digested sludge land requirements were less than for lime stabilized sludge because of the volatile solids reduction during digestion. Truck driving and operation costs were similarly less for digested sludge because of the volatile solids reduction and more concentrated sludge (6 percent versus 4 percent) which would be hauled. Fertilizer credit was less for digested sludge because of the lower amount of dry solids applied to the land. Land credit was based on the amount of sludge applied and was, therefore, less for digested sludge.

The total annual capital and annual operation and maintenance costs for lime stabilization and single stage anaerobic sludge digestion, including land application for a 1 Mgal/day (0.04 m$^3$/s) wastewater treatment plant, are summarized in table 1–21.

Table 1–21.—Comparison of total annual capital and annual O. & M. cost for lime stabilization and anaerobic digestion including land disposal for a 3,785 m$^3$/day plant

| | Lime stabilization | | Anaerobic digestion | |
|---|---|---|---|---|
| | Total annual O. & M. cost | Annual cost per Kkg dry solids | Total annual O. & M. cost | Annual cost per Kkg dry solids |
| Facilities | $23,100 | $59.58 | $42,500 | $109.74 |
| Land application | 17,700 | 45.70 | 8,700 | 19.59 |
| Total annual cost | 40,800 | 105.28 | 51,200 | 129.33 |

## Lime Stabilization by Others

A considerable amount of lime stabilization work has occurred in Connecticut. A number of incinerators have been shut down and replaced by lime stabilization. A total of 27 plants with capacities from 0.3 to 29 mgd (0.01 to 1.27 m$^3$/s) are utilizing lime stabilization either on a full- or part-time basis. The following tabulation and comments for nine plants are typical and summarize the current situation. Lime stabilized sludges are either used as landfill cover or are composted. These methods have been satisfactory. Most of the communities have indicated that they will continue with lime stabilization. Typical plants in Connecticut which are utilizing lime stabilization are shown at the top of the facing page.

| | Design plant size, mgd | Incinerator | | | Lime stabilization | | |
|---|---|---|---|---|---|---|---|
| | | Installed | Used | Hours | Used | Hours | Ultimate disposal |
| Stratford[a] | 11 | Yes | Yes | 24 | Yes | 8 | Landfill cover |
| West Bridgeport[b] | 29 | Yes | Yes | 24 | Yes | 8 | Landfill cover |
| Stamford[c] | 20 | Yes | No | No | N/A | N/A | Lagoon |
| Middletown[d] | 7 | Yes | No | N/A* | Yes | 16 | Landfill cover |
| Willimatic[e] | 5.5 | Yes | No | N/A | Yes | N/A | Land and landfill |
| Glastonburg[f] | 3.2 | Yes | No | N/A | Yes | N/A | Landfill cover |
| Torrington[g] | 7 | Yes | Yes | N/A | Yes | N/A | Landfill |
| Naugatuck[h] | 7 | Yes | No | N/A | Yes | N/A | Landfill cover |
| Enfield[i] | 10 | Yes | Yes | 1/4 of year | Yes | 3/4 of year | Land |

*N/A denotes data not available at the time of writing.
[a]Incinerator abandoned in favor of lime stabilization to pH 12. Composted and used as final cover at landfill.
[b]Stabilized cake used as final cover at landfill.
[c]Centrifuged with lime sludge. Haul away and lagooned.
[d]Previously plagued with odors; now all sludge processed in two shifts, 5 days per week with no odors. Lime stabilized and final cover at landfill.
[e]Began lime stabilization in 1973. Screened sludge and leaf material used on parks as fertilizer and final cover for landfill.
[f]Final cover for landfill and composted with leaves.
[g]Lime stabilization used when incinerator out of service.
[h]Lime stabilized sludge used as final cover at two landfills.
[i]Incineration is used in winter during inclement weather. Lime stabilized sludge stockpiled and spread on corn land during remainder of year.

# LIME STABILIZATION DESIGN EXAMPLES

## Statement of Problem

The problem is to provide lime stabilization facilities for two communities, both of which have existing conventional activated sludge wastewater treatment plants.

The smaller community has existing wastewater treatment facilities capable of treating 4.0 Mgal/day (0.18 $m^3$/s). The facilities consist of screening, grit removal, primary settling, conventional activated sludge aeration, final settling, chlorination, and sludge lagooning. Present flow to the plant is 3.5 million gallons per day (0.15 $m^3$/s); the 20-year projected flow is 4.0 million gallons per day (0.18 $m^3$/s). The plant meets its proposed discharge permit requirements, but the city has been ordered to abandon the sludge lagoons (which are periodically flooded by the receiving stream). Sludge disposal alternatives include the following:

1. Lime stabilization followed by liquid application to farmland.
2. Anaerobic digestion followed by liquid application to farmland.

The larger community has existing wastewater treatment facilities capable of treating 30 million gallons per day (1.31 $m^3$/s). Present flow to the plant is 35 million gallons per day (1.53 $m^3$/s); the 20-year projected flow is 40 million gallons per day (1.75 $m^3$/s). The existing treatment system consists of screening, grit removal, primary settling, conventional activated sludge aeration, final settling, chlorination, aerobic sludge digestion, sludge dewatering, and landfilling of dried sludge solids. The existing treatment scheme will meet proposed permit requirements. As a part of the treatment plant expansion planning and in view of future electric power costs, the following solids handling alternatives were proposed:

1. Lime stabilization followed by pipeline transportation to the land application site.
2. Anaerobic digestion followed by mechanical dewatering and land application.

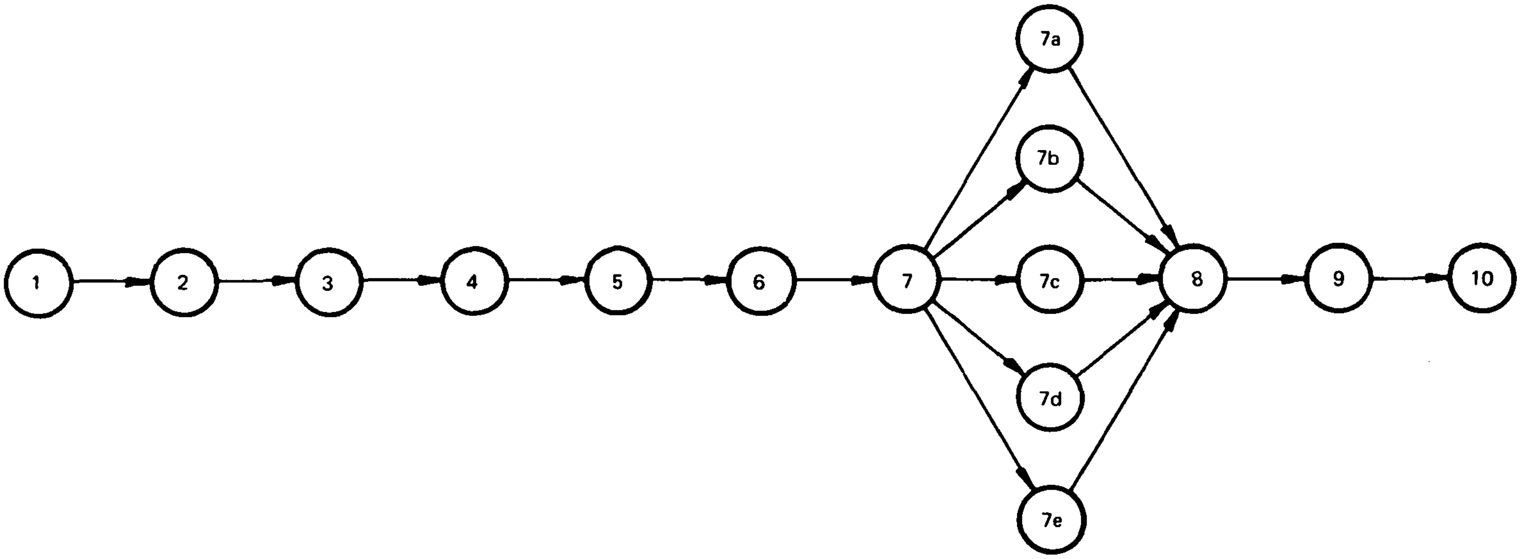
The design logic which will be followed to develop and evaluate the sludge handling alternatives is summarized on figure 1–15.

## Wastewater Characteristics

The wastewater characteristics and removal efficiencies of the various treatment units are required to determine the basis for design of the sludge stabilization and ultimate disposal processes. This information may be acquired from plant records or from sampling programs at the existing facilities. When these data are not available (such as in the case of new wastewater treatment plants for new service areas), assumptions based on sound engineering judgment and previous experience are necessary. For the sake of simplicity, the wastewater characteristics and treatment unit removal efficiencies for the example plants were assumed to be equal. Raw wastewater characteristics for the example plants are given in table 1–22.

## Treatment Unit Efficiencies

Both plants in this example will meet their proposed permit requirements by utilizing the existing treatment




1. Establish regulatory constraints for effluent and sludge disposal
2. *Determine* WWTP *influent loads*
3. Deterimine WWTP unit process scheme
4. Determine raw sludge loads
5. Establish cost effective constraints and sludge solids concentrations for ultimate sludge disposal processes
6. Set sludge thickening requirements
7. Select stabilization alternatives

7a. Develop capital cost
7b. Develop 0 & M requirements and cost
7c. Develop environmental constraints
7d. Evaluate supernatant impact on plant
7e. Evaluate estimated total sludge handling costs
8. Screen alternatives
9. Select final stabilization process
10. Prepare final flow sheets and cost estimates

Figure 1–15.—Process alternative design logic.

Table 1–22.—Raw wastewater characteristics

| Parameter | Concentration (mg/l) |
|---|---|
| $BOD_5$ | 200 |
| Suspended solids | 240 |
| Organic nitrogen | 15 |
| Ammonia nitrogen | 25 |
| Phosphorus | 10 |
| Grease | 100 |

Table 1–23.—Treatment unit efficiencies

| Unit | Parameter | Removal efficiency (percent) |
|---|---|---|
| Primary settling | $BOD_5$ | 30 |
| | SS | 65 |
| Aeration and final settling | $BOD_5$ | 60 |
| | SS | 25 |

processes. Nitrification and phosphorus removal are not required. Removal efficiencies based on percentages of the raw "domestic" wastewater characteristics are presented in table 1–23.

## Sludge Characteristics

The characteristics of sludge discharged to the sludge stabilization facilities may vary considerably depending on the type and amount of industrial waste treated, the sludge origin (which particular treatment unit) and the sludge age. Ideally, samples of sludge would be available for analysis. The assumed sludge characteristics for each example plant are as follows:

| *Sludge type* | *Design percent solids* |
|---|---|
| Thickened raw primary | 7.0 |
| Thickened waste activated | 2.5 |

Thickening facilities for primary and waste activated sludge were assumed to be cost effective for both the 4 and 40 Mgal/d (0.18 and 1.75 $m^3/s$) wastewater treatment plants. Waste activated sludge production was 0.5 pound of volatile solids per pound of $BOD_5$ reduced.

Preliminary studies have indicated that anaerobic sludge digestion will not be adversely affected by the inclusion of thickened waste activated sludge.

The sludge quantities for the 4 Mgal/d (0.18 $m^3/s$) wastewater treatment plant were developed as follows:

Influent $BOD_5$

Influent 4.0 Mgal/d $\times$ 8.34 $\times$ 200 mg/l = 6,672 lb/day (3,033 kg/day)

Primary removal = 6,672 $\times$ 0.3 = 2,002 lb/day (910 kg/day)

$BOD_5$ remaining in settled sewage = 4,670 lb/day (2,123 kg/day)

Influent suspended solids

Influent 4 $\times$ 8.34 $\times$ 240 mg/l = 8,006 lb/day (3,639 kg/day)

Primary removal = 8,006 $\times$ 0.65 = 5,204 lb/day (2,365 kg/day)

Suspended solids remaining in settled sewage = 2,802 lb/day (1,274 kg/day)

Waste activated solids

Biological = 6,672 $\times$ 0.60 $\times$ 0.5 lb VSS/lb $BOD_5$ = 2,002 lb VSS/day

Suspended solids = 8,006 $\times$ 0.25 = 2,002 lb/day

Total biological solids produced = 4,004 lb/day (1,820 kg/day)

Net daily sludge quantities

Primary: 5,204 lb/day (2,360 kg/day) at 7 percent following thickening

$$\frac{5{,}204}{8.34 \times 1.02 \times 0.07} = 8{,}740 \text{ gal/day } (33 \text{ m}^3\text{/day})$$

Waste activated sludge

$$\frac{4{,}004}{0.025 \times 8.34 \times 1.01} = 19{,}014 \text{ gal/day } (72 \text{ m}^3\text{/day})$$

Net sludge produced (5,204 + 4,004) = 9,208 lb solids/day (4,185 kg/day)

Volume = (8,740 + 19,014) = 27,754 gal/day (105 $m^3$day)

Percent solids = 3.9 percent

Design sludge quantities were developed for the 40 Mgal/d (1.75 $m^3/s$) facility in an identical manner. The design sludge quantities are summarized as follows:

| | *4.0 Mgal/d WWTP* | *40 Mgal/d WWTP* |
|---|---|---|
| Primary sludge solids, lb/day | 5,204 | 52,040 |
| Primary sludge volume at 7 percent, gal/day | 8,740 | 87,400 |
| Biological sludge solids, lb/day | 4,004 | 40,040 |
| Biological sludge volume at 2.5 percent, gal/day | 19,014 | 190,140 |
| Total sludge solids, lb/day | 9,208 | 92,080 |
| Combined sludge volume, gal/day | 27,754 | 277,540 |
| Combined sludge percent solids | 3.9 | 3.9 |

For simplicity, the design examples for the 4 and 40 Mgal/d (0.18 and 1.75 $m^3/s$) treatment plants will be presented separately. Each example will include the design basis for each alternative stabilization and ultimate disposal process, final sludge volumes, capital and annual operation and maintenance costs.

## Process Alternatives—4 Mgal/d (0.18 $m^3/s$) WWTP

As previously discussed, process alternatives for the 4 Mgal/d wastewater treatment plant will be as follows:

1. Lime stabilization followed by liquid application to farmland.
2. Anaerobic digestion followed by liquid application to farmland.

### *Lime Stabilization*

A flow diagram for the proposed lime stabilization facilities is shown on figure 1–16. Significant process equipment includes a bulk lime storage bin for pebble quicklime, auger, lime slaker and feed slurry tank, sludge mixing and thickening tank with a mechanical mixer, sludge grinder, all weather treatment building, electrical and instrumentation, interconnecting piping and transfer pumps, and a sludge holding lagoon with 60 days detention time. The basis for design is as follows:

| | |
|---|---|
| Total sludge solids | 9,208 lb/day (4,185 kg/day) |
| Sludge volume | 27,754 gal/day (105 $m^3$/day) |
| Raw sludge percent solids | 3.9 |
| Overall lime dosage required per unit dry solids, as 100 percent $Ca(OH)_2$ | 0.20 lb/lb |
| Daily lime requirement as $Ca(OH)_2$ | 1,826 lb/day (830 kg/day) |
| Treatment period | 6 hrs/day |
| Bulk lime storage bin volume minimum | 1,000 $ft^3$ (28 $m^3$) |
| Bulk lime storage bin detention time | 34 days |
| Lime slaker and slurry tank capacity (2 ea) | 200–300 lb CaO/hr (91–136 kg/hr) |
| Influent sludge grinder capacity (spared) | 200 gal/min (12.6 l/s) |
| Sludge mixing tank volume | 25,000 gal (95 $m^3$) |
| Sludge mixing tank dimensions | 18 ft $\times$ 18 ft $\times$ 10 ft SWD (5.5 m $\times$ 5.5 m $\times$ 3 m) |
| Sludge mixer horsepower | 15 HP (11.2 kW) |
| Sludge mixer turbine diameter | 53 in (135 cm) |
| Turbine speed | 68 rpm |
| Sludge transfer pump capacity (spared) | 400 gal/min (25.2 l/s) |
| Treated sludge volume | 24,050 gal (91 $m^3$) |
| Treated sludge percent solids | 4.5 |
| Sludge holding lagoon total volume (4 cells) | 240,000 $ft^3$ (6,800 $m^3$) |
| Sludge holding lagoon maximum detention time | 60 days |
| Treatment building floor area | 250 $ft^2$ (23.2 $m^2$) |
| Treatment building construction | brick and block |
| Instrumentation | pH record treated sludge volume |

With the exception of the lime storage bin detention time and pump capacities, the reasons for selecting the particular design quantities have been discussed in previous sections. Lime storage bin capacity was based on a minimum detention time of 30 days to allow capacity for a standard 20-ton (18 Mg) lime shipment. The pump capacity was based on convenient transfer times between units.

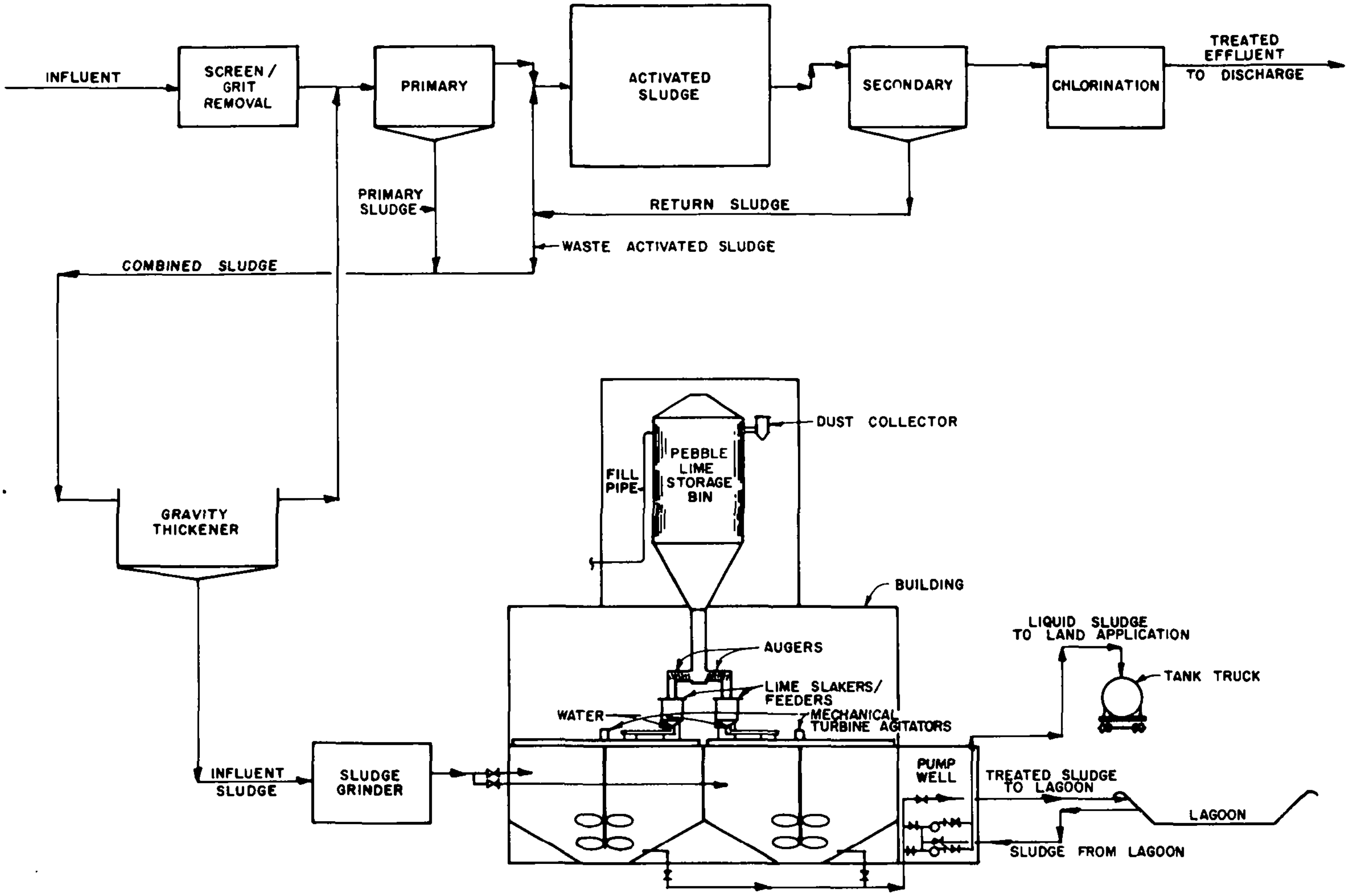



Figure 1–16.—4 Mgal/d (0.18 $m^3/s$) lime stabilization/truck haul and land application.

Capital costs for the lime stabilization facilities were based on January 1, 1978, bid date and have been summarized in table 1–24.

Lime stabilization operation assumed one man, 8 hours per day, 365 days per year, at $6.50 per hour, including overhead. Maintenance labor was assumed to be 156 hours per year labor at $6.50 per hour and $2,400 per year for maintenance materials. The total quantity of 85 percent CaO quicklime required was 297 tons (269 Mg) per year at $40 per ton ($44/Mg).

The total annual cost for lime stabilization, excluding land application of treated sludge, has been calculated as follows and is summarized in table 1–25.

### Lime Stabilization Operating Costs

Labor: 8hr/day × 365 day/yr × $6.50/hr = $18,980 say $19,000

Table 1–24.—Capital costs of lime stabilization facilities for a 4 Mgal/d wastewater treatment plant

| | |
|---|---|
| Site work, earthwork, yard piping | $26,000 |
| Lime storage bin and feeders | 84,000 |
| Treatment tank, pumps, sludge grinders, and building structure | 142,000 |
| Electrical and instrumentation | 29,000 |
| Sludge holding lagoon | 54,000 |
| Subtotal construction cost | 335,000 |
| Engineering | 36,000 |
| Total capital cost | 371,000 |
| Amortized cost at 30 yrs., 7 percent int. (CRF = 0.081) | 30,100 |
| Annual capital cost per ton dry solids | 17.91 |

Table 1–25.—Total annual cost for lime stabilization excluding land disposal for a 4 Mgal/d plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Operating labor | $19,000 | $11.31 |
| Maintenance labor and materials | 3,400 | 2.02 |
| Lime | 12,000 | 7.14 |
| Laboratory | 1,500 | 0.89 |
| Capital | 30,100 | 17.91 |
| Total annual cost | 66,000 | 39.27 |

Maintenance labor: 156 hr/yr × \$6.50 = \$1,014 say \$1,000
Maintenance materials: \$2,400/yr lump sum
Lime primary: 5,204 lb/day × 0.12 lb $Ca(OH)_2$/lb = 624 lb/day (283 kg/day)
Waste activated: 4,004 lb/day × 0.3 lb $Ca(OH)_2$/lb = 1,201 lb/day (545 kg/day)
Total lime = (624 + 1,201) = 1,825lb $Ca(OH)_2$/day (828 kg/day)
1,825 lb/day/0.85) × 56/74 = 1,625 lb/day CaO (737 kg/day)
1,625 × 365/2,000 = 297 ton/yr (269 Mg/yr)
say 300 ton/yr × \$40/ton = \$12,000/yr
Laboratory: \$1,500/yr lump sum
Capital: \$371,000 × 0.081 = \$30,100/yr

Both the lime stabilization and anaerobic digestion alternatives were assumed to utilize land application of treated sludge as a liquid hauled by truck. The capital cost per sludge hauling vehicle was assumed to be \$35,000, which was depreciated on a straight-line basis over a 5-year period.

The assumed hauling distance was 3 to 5 miles (5–8 km), round trip. Hauling time assumed 10 minutes to fill, 15 minutes to empty, and 10 minutes driving, or a total of 35 minutes per round trip. The truck volume was assumed to be 1,500 gallons (5.7 $m^3$) per load. The cost of truck operations, excluding the driver and depreciation, was assumed to be \$8.50 per operating hour. The truck driver labor rate was assumed to be \$6.50 per hour, including overhead.

Truck operation time was based on hauling on a 5-day per week basis, approximately 10 months per year, which results in the assumed 215 hauling days per year. The average volume hauled is 40,800 gallons (154.4 $m^3$) per day. Two trucks were assumed to be required, with a combined total of 28 loads per day.

Although it may be possible to obtain the use of farmland at no cost, e.g., on a voluntary basis, the land application economic analysis assumed that land would be purchased at a cost of \$750 per acre (\$1850/ha). Sludge application rates were assumed to be 10 dry tons per acre per year. Land costs were amortized at 7 percent interest over a 30-year period.

To offset the land cost, a fertilizer credit of \$7.30 per ton (\$8.05 Mg) of dry sludge solids was assumed. This rate was arbitrarily assumed to be 50 percent of the value published by Brown[11] based on medium fertilizer market value and low fertilizer content. The reduction was made to reflect resistance to accepting sludge as fertilizer. The land cost was further offset by assuming a return of \$50 per acre (\$123/ha), either as profit after farming expenses or as the rental value of the land.

Capital and annual operation and maintenance costs for land application of lime stabilized sludge were calculated as follows and have been summarized in table 1–26.

**Lime Stabilization Land Application Costs**

Land: 9,208 lb solids/day × 365 days/2,000 lb/ton = 1,681 ton/yr (1,525 Mg/yr)
1,681 ton/yr/10 ton/acre = 168 acres (68.0 ha) say 200 (80.9 ha)
200 acres × \$750/acre = \$150,000
\$150,000 × 0.081 = \$12,150/yr say \$12,200
Truck depreciation: \$35,000 × 2 = \$70,000 capital
\$70,000/5 yrs = \$14,000/yr
Truck driver: 40,800 gal/day/2,571 gal/truck/hr = 15.9 hr/day
say 2 trucks at 8 hr/day
\$6.50 × 2 men × 8 hr/day = \$104/day
\$104 × 215 = \$22,360 say \$22,400/yr
Truck operation: 2 trucks × 8 hr/day × \$8.50/hr = \$136.00/day
\$136.00 × 215 = \$29,240 say \$29,200/yr
Laboratory: \$1,500/yr lump sum
Fertilizer credit: 1,681 ton/yr × \$7.30/ton = \$12,271 say \$12,300/yr
Land credit: 168 acres × \$50/acre = \$8,400/yr

Table 1–26.—Annual cost for land application of lime stabilized sludge for a 4 Mgal/d plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Amortized cost of land | $12,200 | $7.26 |
| Truck depreciation | 14,000 | 8.33 |
| Truck driver | 22,400 | 13.33 |
| Truck operation | 29,200 | 17.38 |
| Laboratory | 1,500 | 0.89 |
| Fertilizer credit | (12,300) | (7.30) |
| Land credit | (8,400) | (5.00) |
| Total annual cost | 58,600 | 34.89 |

## *Anaerobic Digestion*

A flow diagram for the proposed anaerobic sludge digestion facilities is shown on figure 1–17. Two-stage anaerobic digestion was assumed with stabilized sludge being hauled to farmland. Sludge storage was allowed in the digester design and no lagoon was included. The basis for design for the anaerobic digesters for the 4 Mgal/d (.18 $m^3$/s) treatment plant was as follows:

First Stage

| | |
|---|---|
| Feed solids loading | 9,208 lb/day (4,185 kg/day) |
| Feed volume | 27,754 gal/day (105 $m^3$/s) |
| Feed percent solids | 3.9 |
| Feed percent volatile solids | 65 |
| Digester dimensions | 60 ft × 25 ft SWD (18.3 m × 7.6 m) |
| Digester volume | 529,000 gal (2,002 $m^3$) |
| Mixers | 2 ea at 3,500 gpm (221 l/s) |
| Hydraulic detention time | 19 days |
| Loading rate | 0.085 lb/VSS/$ft^3$/day (1.36 kg/$m^3$/day) |
| Digester bulk temperature | 95°F (35°C) |
| Average feed temperature | 55°F (13°C) |
| Volatile solids reduction | 50 percent |
| Overall total solids reduction | 32 percent |
| Sludge heaters | 3 ea at 500,000 Btu/hr (14,650 W) |

Second Stage

| | |
|---|---|
| Digester dimensions | 60 ft × 25 ft SWD (18.3 m × 7.6 m) |
| Digester volume | 529,000 gal (2,002 $m^3$) |
| Hydraulic detention time | 19 days |
| Digester gas production | 10 $ft^3$/lb VSS (0.6 $m^3$/kg) feed |
| Digester gas heat value | 500 Btu/$ft^3$ (18,625 kJ/$m^3$) |
| Digested sludge dry solids production | 6,261 lb/day (2,846 kg/day) |
| Digested sludge percent solids | 6.5 percent |
| Sludge recirculation pumps (2 ea) | 500 gpm ea (31.5 l/s) |

Design conditions were based on the criteria enumerated in Ten States' Standards[28] and assumed installation in the Midwestern United States.

Capital cost for the anaerobic sludge digestion facilities, including the control building, structures, floating cover, heat exchanger, gas safety equipment, pumps, and interconnecting piping, assuming January 1, 1978, bid date, and engineering, legal, and administrative costs is summarized in table 1–27.

Digester operation assumed one man, 3 hours per day, 365 days per year at $6.50 per hour, including overhead. Maintenance labor and material assumed 416

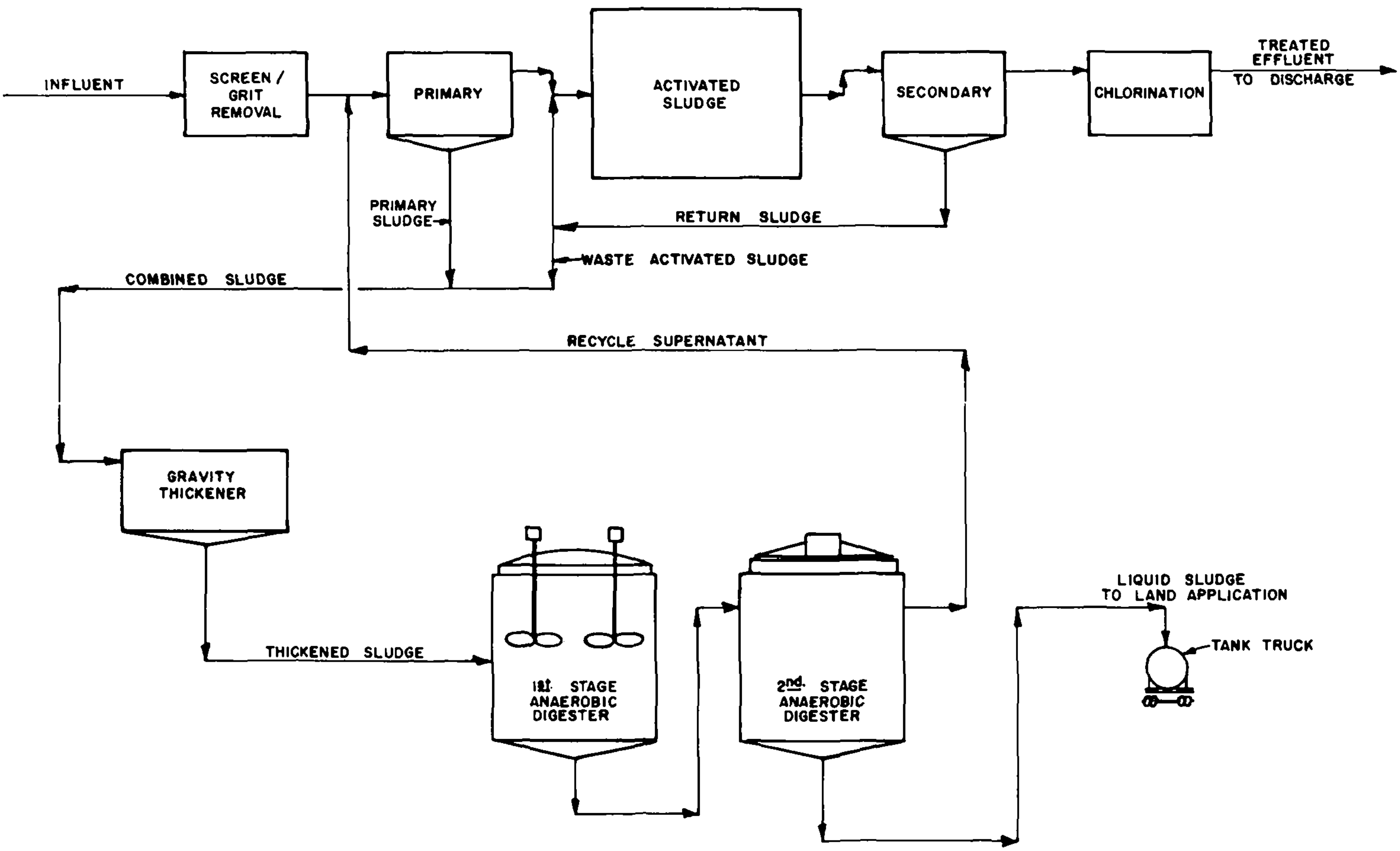



Figure 1–17.—4 Mgal/d (0.18 $m^3$/s) anaerobic digestion/truck haul and land application.

Table 1–27.—Capital cost of two-stage anaerobic digestion facilities for a 4 Mgal/d wastewater treatment plant

| | |
|---|---|
| Site work, earthwork, yard piping and pumps | $151,000 |
| Digesters | 675,000 |
| Control building | 251,000 |
| Electrical and instrumentation | 125,000 |
| Subtotal construction cost | 1,202,000 |
| Engineering | 107,000 |
| Total capital cost | 1,309,000 |
| Amortized cost at 30 yrs., 7 percent int. (CRF = 0.081) | 106,000 |
| Annual capital cost per unit feed dry solids | 63.08 |

hours per year at $6.50 per hour and $7,000 per year for maintenance materials.

The cost of anaerobic digester operation was offset by assuming a value of $2.70 per million Btu ($2.56 per million kJ) for all digester gas produced above the net digester heat requirements.

The total annual cost for anaerobic sludge digestion, excluding land application was calculated as follows and has been summarized in table 1–28.

### Anaerobic Digester O & M Cost

Operator labor: 3 hr/day $\times$ 365 day/yr $\times$ $6.50/hr = $7,118/yr say $7,100/yr

Maintenance labor: 416 hr/yr $\times$ $6.50/hr = $2,704 say $2,700/yr

Maintenance materials: $7,000/yr lump sum

Laboratory: $1,500/yr lump sum

Capital $1,309,000 $\times$ 0.081 = $106,000

Fuel credit: 9,208 lb $\times$ 0.65 = 5,985 lb VSS feed/day

5,985 lb $\times$ 10 cf/lb VSS = 59,850 cf/day (1,695 $m^3$/day) gas

59,850 $ft^3$ $\times$ 500 Btu/$ft^3$ = $29.9 \times 10^6$ Btu/day ($31.6 \times 10^6$ kJ/day)

475,000 Btu/hr $\times$ 24 hr/day/0.5 eff = $22.8 \times 10^6$ Btu/day ($24.1 \times 10^6$ kJ/day) required for digester heat

$29.9 \times 10^6 - 22.8 \times 10^6 = 7.1 \times 10^6$ Btu/day ($7.5 \times 10^6$ kJ/day) excess gas

$7.1 \times 10^6 \times \$2.70 \times 10^{-6} \times 365 = \$6,997$ say $7,000/yr

Land application costs were developed for the anaerobic digestion alternative in a manner similar to that previously described for lime stabilization. Anaerobically digested sludge land requirements were less than for lime stabilized sludge because of the volatile solids reduction during digestion. Truck driving and operation costs were similarly less for digested sludge because of the volatile solids reduction and more concentrated sludge (6.5 percent versus 4.5 percent) which would be hauled. The total fertilizer credit was based on $7.30 per ton ($8.05/Mg) of dry solids, but was lower because of the lower amount of dry solids applied to the land. The total land credit was less because land requirements were based on the total amount of sludge solids applied. Land application costs for the anaerobic digestion alternative were calculated in a manner similar to those for the lime stabilization alternative and are summarized in table 1–29.

The total annual capital and annual operation and maintenance costs for lime stabilization and two-stage anaerobic sludge digestion, including land application for a 4 Mgal/d (0.18 $m^3$/s) wastewater treatment plant, are summarized in table 1–30.

Table 1–28.—Total annual cost for two-stage anaerobic sludge digestion excluding land disposal for a 4 Mgal/d plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Operating labor | $7,100 | $4.23 |
| Maintenance labor and materials | 9,700 | 5.77 |
| Laboratory | 1,500 | 0.89 |
| Capital | 106,000 | 63.08 |
| Fuel credit | (7,000) | (4.16) |
| Total annual cost | 117,300 | 69.81 |

## Process alternatives—40 Mgal/d (1.75 $m^3$/s) WWTP

As previously discussed, process alternatives for the 40 Mgal/d wastewater treatment plant will be as follows:

1. Lime stabilization followed by pipeline transportation to the land application site.
2. Anaerobic digestion followed by mechanical dewatering and land application.

Table 1–29.—Annual cost for land application of anaerobically digested sludges for a 4 Mgal/d plant

| Item | Total annual cost | Annual cost per ton solids |
|---|---|---|
| Amortized cost of land | $8,200 | $4.88 |
| Truck depreciation | 7,000 | 4.17 |
| Truck driver | 11,200 | 6.66 |
| Truck operation | 14,600 | 8.69 |
| Laboratory | 1,500 | 0.89 |
| Fertilizer credit | (8,300) | (4.94) |
| Land credit | (5,700) | (3.39) |
| Total annual cost | 28,500 | 16.96 |

Table 1–30.—Comparison of total annual capital and annual O. & M. cost for lime stabilization and anaerobic digestion including land disposal for a 4 Mgal/d plant

| | Lime stabilization | | Anaerobic digestion | |
|---|---|---|---|---|
| | Total annual O. & M. cost | Annual cost per ton dry solids | Total annual O. & M. cost | Annual cost per ton dry solids |
| Facilities | | | | |
| Amortized capital | $30,100 | $17.91 | $106,000 | $63.08 |
| Operating labor | 19,000 | 11.31 | 7,100 | 4.23 |
| Maintenance labor and materials | 3,400 | 2.02 | 9,700 | 5.77 |
| Lime | 12,000 | 7.14 | | |
| Laboratory | 1,500 | 0.89 | 1,500 | 0.89 |
| Fuel credit | N/A | N/A | (7,000) | (4.16) |
| Subtotal facilities | 66,000 | 39.27 | 117,300 | 69.81 |
| Land application | | | | |
| Amortized cost of land | 12,200 | 7.26 | 8,200 | 4.88 |
| Truck depreciation | 14,000 | 8.33 | 7,000 | 4.17 |
| Truck drivers | 22,400 | 13.33 | 11,200 | 6.66 |
| Truck operations | 29,200 | 17.38 | 14,600 | 8.69 |
| Laboratory | 1,500 | 0.89 | 1,500 | 0.89 |
| Fertilizer credit | (12,300) | (7.30) | (8,300) | (4.94) |
| Land credit | (8,400) | (5.00) | (5,700) | (3.39) |
| Subtotal land application | 58,600 | 34.89 | 28,500 | 16.96 |
| Total annual cost facilities and land application | 124,600 | 74.16 | 145,800 | 86.77 |

The design logic which will be followed to develop and evaluate the sludge handling alternatives has previously been summarized on figure 14. Wastewater characteristics, treatment unit efficiencies, and sludge characteristics have also been previously summarized.

## *Lime Stabilization*

A flow diagram for the proposed lime stabilization facilities is shown on figure 1–18. Significant process equipment includes a bulk lime storage bin for pebble quicklime, augers, lime slakers and feed slurry tanks, sludge mixing tanks, sludge thickeners, sludge grinders, all weather treatment building, electrical and instrumentation, interconnecting piping, and sludge pump stations.

The sludge pipeline was assumed to be 10 miles (1.6 km) long with two intermediate pump stations. One land application farm site was assumed. A sludge storage lagoon with 60-days holding capacity was provided at the land application site.

The basis for design is as follows:

| | |
|---|---|
| Total sludge solids | 92,080 lb/day (41,855 kg/day) |
| Sludge volume | 227,540 gal/day (861.2 $m^3$/day) |
| Raw sludge percent solids | 3.9 |
| Overall lime dosage required per unit dry solids as 100 percent $Ca(OH)_2$ | 0.20 lb/lb dry solids |
| Daily lime requirement as $Ca(OH)_2$ | 18,250 lb/day (8,295 kg/day) |
| Treatment period | 24 hrs/day |
| Bulk lime storage bin volume minimum | 2 ea 4,260 $ft^3$ (120.6 $m^3$) |
| Bulk lime storage bin detention time | 30 days |
| Lime slaker & slurry tank capacity (2 ea) | 500–750 CaO/hr |
| Influent sludge grinder max. capacity | 2 ea 200 gpm (12.6 l/s) |
| Sludge mixing tank volume at 1 hr detention time (2 ea) | 12,000 gal (45.4 $m^3$) |
| Sludge mixing tank dimensions | 10 ft × 10 ft SWD (3 m × 3 m) |
| Sludge mixer horsepower (2 ea) | 10 HP (7.5 kW) |
| Sludge mixer turbine diameter | 5 ft (1.5 m) |
| Turbine speed | 45 rpm |
| Sludge thickener dimensions (2 ea) | 65 ft dia. × 12 ft SWD (19.8 m × 3.7 m) |
| Thickened sludge volume | 240,500 gal/day (910.3 $m^3$/day) |
| Thickened sludge percent solids | 4.5 |
| Sludge transfer pump capacity (2 ea) | 250 gpm at 200 psi (15.8 l/s at 14.1 kg $km^2$) |
| Intermediate pump station pumps | 4 ea 250 gpm at 200 psi (15.8 l/s at 14.1 kg/$cm^2$) |
| Treatment building floor area | 600 $ft^2$ (55.7 $m^2$) |
| Treatment building construction | brick and block |
| Instrumentation | pH record/control raw sludge volume treated sludge volume pipeline pressure control |
| Lagoon volume at application site | 10,000,000 gal (37,850 $m^3$) (20 cells) |
| Pipeline length | 53,000 ft (16,154 m) |
| Pipeline diameter | 6 in (15 cm) |
| Pipeline working pressure | 200–250 psig (14.1–17.6 kg/$cm^2$) |
| Land application trucks | 12 at 1,500 gal (5.7 $m^3$) ea |

The reasons for selecting the particular design quantities have been discussed in previous sections. Sludge pump capacities were selected to permit reasonable pipeline pressure drops and velocities. The sludge lagoon was divided into several cells to permit convenient withdrawal of all sludge and to prevent solids accumulation.

*Capital costs for the lime stabilization facilities,* based on January 1, 1978, bid date, excluding final sludge pumping, pipeline, application trucks, lagoon, and land, are summarized in table 1–31.

Lime stabilization operation assumed two men, three shifts per day, 365 days per year at $6.50 per hour, including overhead. Maintenance labor was assumed to be 1,664 hours per year at $6.50 per hour and $7,500 per year for maintenance materials. The total quantity of 85 percent CaO quicklime required was 2,966 tons (2691 Mg) per year at $40 per ton ($44/Mg).

The total annual cost for lime stabilization, excluding land application of treated sludge, was calculated in a manner to that previously shown on the 4 Mgal/d (0.18 $m^3$/s) example and has been summarized in table 1–32.

Ultimate sludge disposal was assumed to be as a liquid on farmland with truck spreading. The total land

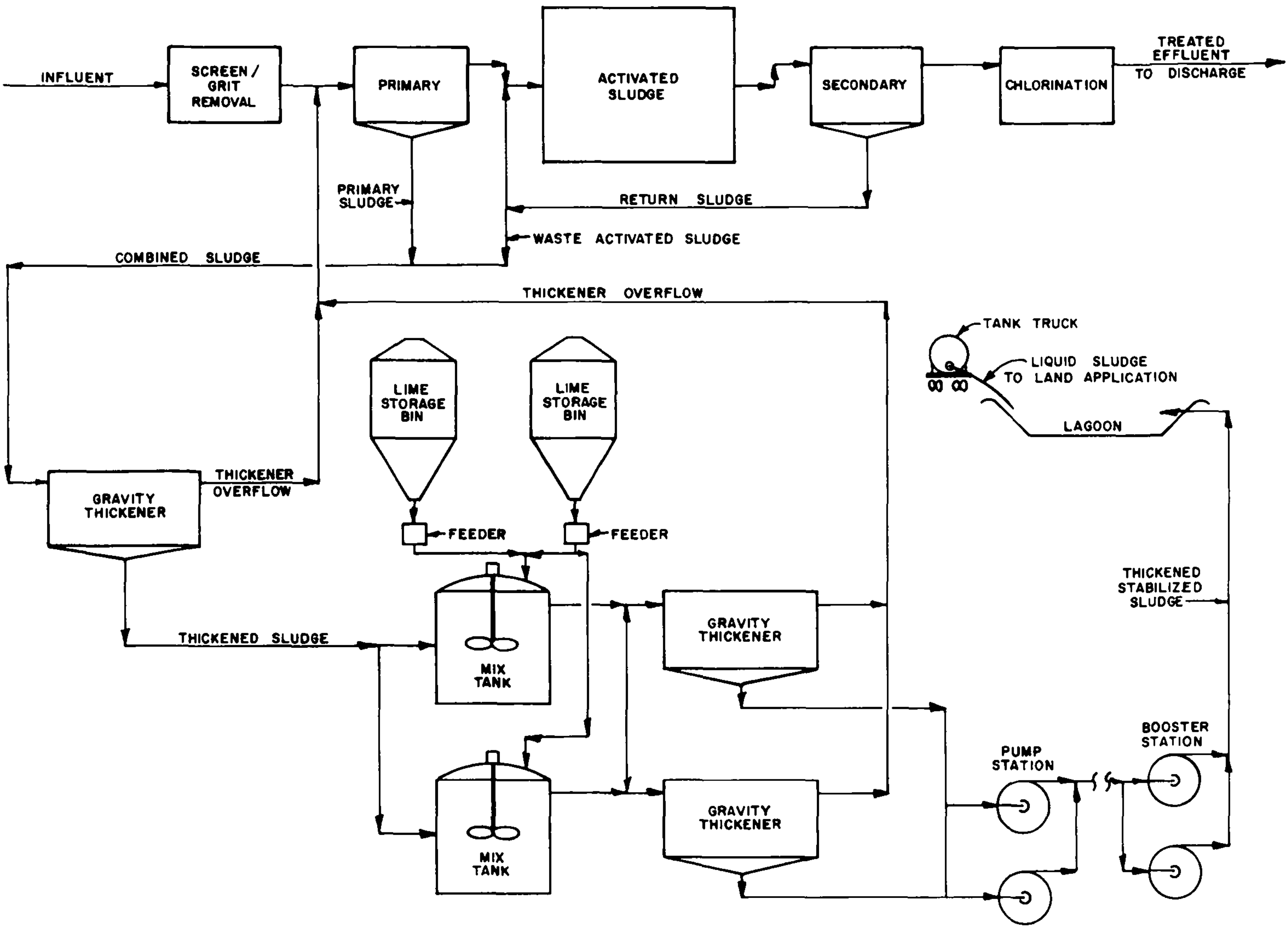



Figure 1–18.—40 Mgal/d (1.75 $m^3$/s) lime stabilization/pipeline transport and land application.

Table 1–31.—Capital cost of lime stabilization facilities for a 40 Mgal/d wastewater treatment plant

| | |
|---|---|
| Site work, earthwork and yard piping | $95,000 |
| Lime storage, slakers, and feed | 106,000 |
| Lime treatment tanks, mixers, grinders and building | 155,000 |
| Sludge thickeners | 529,000 |
| Electrical and instrumentation | 102,000 |
| Subtotal construction cost | 987,000 |
| Engineering | 90,000 |
| Total capital cost | 1,077,000 |
| Amortized cost at 30 yrs., 7% int. (CFR = 0.081) | 87,200 |
| Annual capital cost per unit feed dry solids | 5.19 |

Table 1–32.—Total annual cost for lime stabilization excluding land disposal for a 40 Mgal/d plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Operating labor | $114,000 | $6.78 |
| Maintenance labor and materials | 18,300 | 1.09 |
| Lime | 118,600 | 7.06 |
| Power | 2,000 | 0.12 |
| Laboratory | 4,500 | 0.27 |
| Total annual cost | 257,400 | 15.32 |

spreading operation will require 2,000 acres (810 ha). Land cost was assumed to be $1,250 per acre ($3088/ha) to reflect the more urban setting than the 4 Mgal/d (0.18 $m^3/s$) case. The capital cost per sludge hauling vehicle was assumed to be $35,000, with 12 being required. The vehicles were depreciated over a 7-year period. The sludge holding lagoon was located at the farm site and was sized to hold 60-days sludge production. The lagoon was partitioned into 500,000 gallon (1,892 $m^3$) cells to permit access and efficient utilization of the storage volume.

The assumed hauling time was 10 minutes to fill, 20 minutes to haul, empty and return, for a total of 30 minutes per round trip. The truck volume was assumed to be 1,500 gallons (5.7 $m^3$) per load. The cost of truck operations, excluding the driver and depreciation, was assumed to be $8.50 per operating hour. The truck driver labor rate was assumed to be $6.50 per hour, including overhead.

Truck operating time was based on hauling on a 215-day-per-year schedule, 12 hours per day.

To offset the land cost, a fertilizer credit of $7.30 per ton ($8.05/Mg) of dry sludge solids was assumed. This rate was arbitrarily assumed to be 50 percent of the value published by Brown[11] based on medium fertilizer market value and low fertilizer content. The reduction was made to reflect resistance to accepting sludge as fertilizer. The land cost was further offset by assuming a return of $50 per acre ($124/ha), either as profit after farming expenses, or as the rental value of the land.

Easements for the sludge pipeline were assumed to cost $2.50 per foot ($8.20/m). Two intermediate booster stations were provided to maintain a reasonable pressure profile along the line. Progressive cavity pumps were used for both the treatment plant and intermediate pump stations. Allowance was assumed to permit regular cleaning of the line by utilizing pipeline "pigs."

Table 1–33.—Capital cost of lime stabilization land application facilities for a 40 Mgal/d wastewater treatment plant

| | |
|---|---|
| Site work, earthwork | $17,000 |
| Sludge transfer pumps | 45,000 |
| Sludge pipeline | 675,000 |
| Booster station | 104,000 |
| Sludge lagoon | 569,000 |
| Electrical and instrumentation | 19,000 |
| Subtotal construction cost | 1,429,000 |
| Engineering | 124,000 |
| Total capital cost pipeline, pump stations and lagoon | 1,553,000 |
| Amortized cost at 30 yrs., 7 percent int. (CFR = 0.081) | 125,800 |
| Annual capital cost per unit feed dry solids | 7.49 |

Table 1–34.—Annual cost for transportation and land application of lime stabilized sludge for a 40 Mgal/d plant

| Item | Capital cost | Total annual cost | Annual cost per ton dry solids |
|---|---|---|---|
| Land | $2,500,000 | $202,500 | $12.05 |
| Easements | 132,000 | 10,700 | 0.64 |
| Pipeline, pump stations and lagoon | 1,553,000 | 125,800 | 7.49 |
| Truck depreciation | 420,000 | 60,000 | 3.57 |
| Truck drivers | | 201,200 | 11.97 |
| Truck operation | | 263,200 | 15.66 |
| Power | | 35,000 | 2.08 |
| Pipeline operation and maintenance | | 17,000 | 1.01 |
| Laboratory | | 4,500 | 0.27 |
| Fertilizer credit | | (122,700) | (7.30) |
| Land credit | | (84,000) | (5.00) |
| Total annual cost | 4,605,000 | 713,200 | 42.44 |

Capital costs for the lime stabilization land application site, based on January 1, 1978, bid date, have been summarized in table 1–33.

Annual operation and maintenance costs for transportation and land application of lime stabilized sludge were calculated in a manner similar to that previously summarized and have been shown in table 1–34.

## *Anaerobic Digestion*

A flow diagram for the proposed anaerobic digestion/vacuum filtration alternative is shown on figure 1–19. Significant process equipment includes two-stage standard rate anaerobic sludge digestion, bulk lime and ferric chloride storage, lime slakers, vacuum filtration, sludge conveyors, and sludge storage bin. All facilities were assumed to be housed in an all weather brick-block type building and included all electrical, instrumentation, interconnecting piping, and sludge pumps. The existing sludge dewatering equipment was assumed not to be capable of functioning over the project life and was replaced. Similarly, the existing filter building and chemical feed facilities were replaced.

Design data for the anaerobic digester alternative are as follows:

| | |
|---|---|
| Primary digesters | 3 ea 110 ft × 30 ft (33.5 m × 9.1 m) SWD |
| Secondary digesters | 3 ea 110 ft × 30 ft (33.5 m × 9.1 m) SWD |
| Vacuum filtration | 3 ea at 400 $ft^2$ ea (37.2 $m^2$) |
| Vacuum filter loading rate | 3.5 lb dry solids/$ft^2$/hr (17.1 kg/$m^2$/hr) |
| Lime storage bin | 1 ea 4,000 $ft^3$ (113.3 $m^3$) |

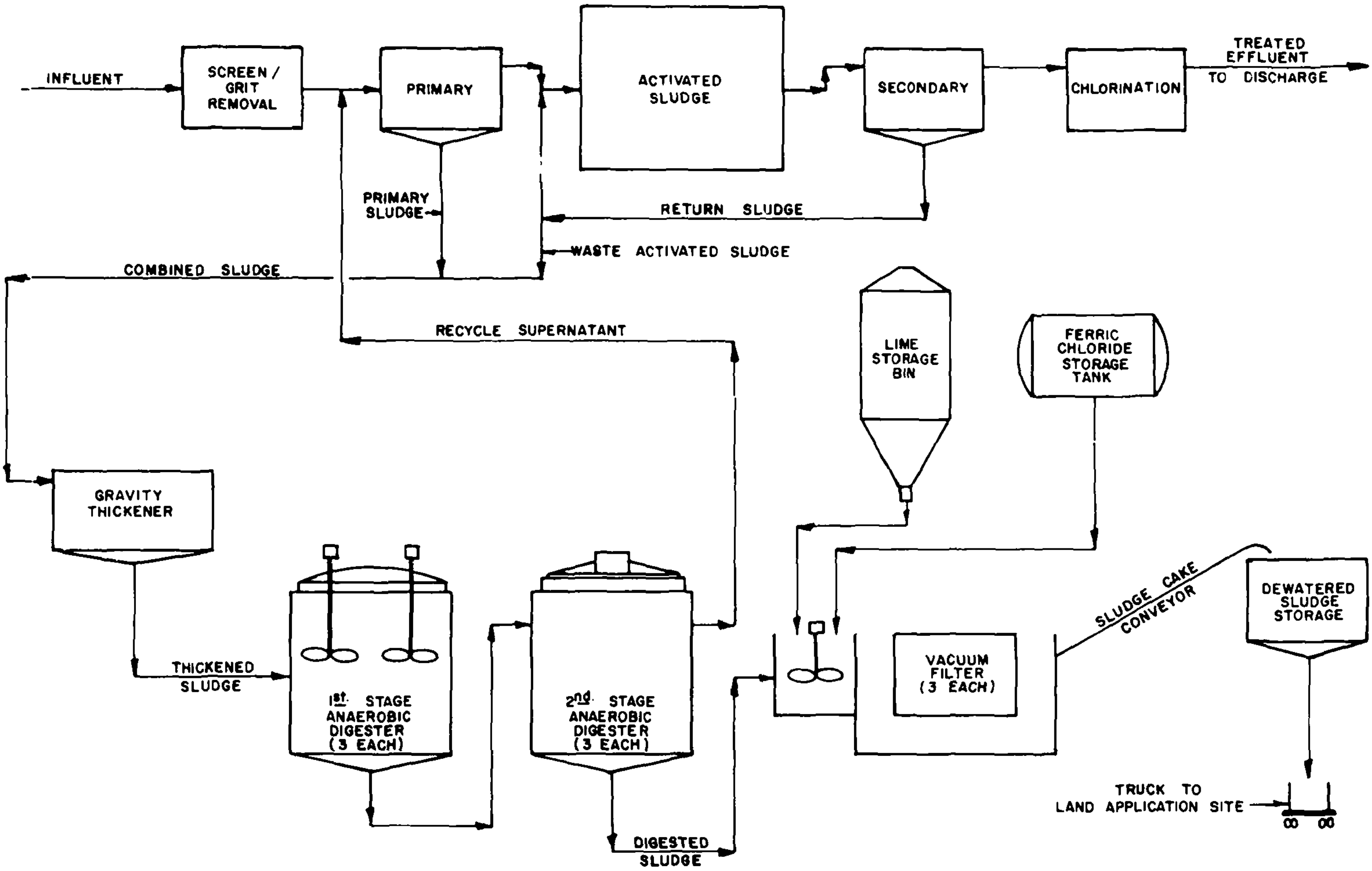



Figure 1–19.—40 Mgal/d (1.75 $m^3$/s) anaerobic digestion/vacuum filtration and land application.

| | |
|---|---|
| Lime slaker/feeders | 3 at 250–500 lb CaO/hr (113.6–227.3 kg/hr) |
| Ferric chloride storage tanks | 2 ea at 5,000 gal ea (18.9 $m^3$) |
| Dewatered sludge storage bin | 1 ea at 2,000 $ft^3$ (56.6 m) |
| Filter building | 3,000 $ft^2$ (278.7 $m^2$) w/basement |
| Digester loading—1st stage | 0.07 lb VSS/$ft^3$/day (1.1 kg/$m^3$/day) |
| Hydraulic detention time—1st stage | 23 days |
| Digester gas production | 10 $ft^3$/lb VSS feed (0.6 $m^3$/kg) |
| Digester gas heat value | 500 Btu/$ft^3$ (18,625 kJ/$m^3$) |
| Volatile solids reduction | 50 percent |
| Overall solids reduction | 32 percent |
| Sludge mixers | 4 at 5,000 gpm ea (315.4 l/s) |
| Digester heat requirement (primary only) | $22.7 \times 10^7$ Btu/day ($2.8 \times 10^6$ W) |
| Gas production | $30.0 \times 10^7$ Btu/day ($3.7 \times 10^6$ W) |
| Net gas available | $7.3 \times 10^7$ Btu/day ($0.9 \times 10^6$ W) |

Design conditions were based on the criteria enumerated in Ten States' Standards[28] and assumed installation in the midwest.

Annual capital costs operation and maintenance for the anaerobic digestion facilities were based on the January 1, 1978 bid date and have been summarized in table 1–29. Capital costs included the digesters, control buildings, covers, heat exchangers, gas safety equipment, interconnecting piping, engineering, legal and administrative costs. Capital costs are summarized in table 1–35.

Digester operation assumed one man, two shifts per day, 365 days per year, at $6.50, including overhead. Maintenance labor and material assumed 4,160 hours per year at $6.50 per hour and $30,000 per year for maintenance materials.

The cost of anaerobic digester operation was offset by assuming a value of $2.70 per million Btu ($2.56 per million kJ) for all digester gas produced above the net digester heat requirement. Total annual operation and maintenance cost for the digestion facilities is summarized in table 1–36.

Table 1–35.—Capital cost of two-stage anaerobic digestion facilities for a 40 Mgal/d (1.75 $m^3/s$) wastewater treatment plant

| | |
|---|---|
| Site work, earthwork, yard piping | $688,000 |
| Digesters and control building | 7,222,000 |
| Pumping | 35,000 |
| Electrical and instrumentation | 745,000 |
| Subtotal construction cost | 8,690,000 |
| Engineering | 649,000 |
| Total capital cost | 9,339,000 |
| Amortized cost at 30 yrs., 7 percent int. (CFR = 0.081) | 756,500 |
| Annual capital cost per unit feed dry solids | 45.02 |

Table 1–36.—Total annual cost for two-stage anaerobic sludge digestion excluding vacuum filtration and land disposal for a 40 Mgal/d (1.75 $m^3/s$) plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Operating labor | $38,000 | $2.26 |
| Maintenance labor and materials | 57,000 | 3.39 |
| Laboratory | 6,000 | 0.36 |
| Capital | 756,500 | 45.02 |
| Fuel credit | (71,900) | (4.28) |
| Total annual cost | 785,600 | 46.75 |

Capital costs for the filtration facilities are summarized in table 1–37

Vacuum filtration costs were estimated as summarized in table 1–38.

Land application costs were calculated based on hauling 20 miles (32 km) round trip. A sludge transfer site was assumed to be located at the land application site. Sludge transfer trucks were assumed to be equipped with 8 $yd^3$ (6.1 $m^3$) dump beds. A total of four trucks were required, operating 8 hours per day, 215 days per year. The loader and land spreading vehicle were assumed to operate 8 hours per day. Land application vehicles were assumed to have 17 $yd^3$ (13 $m^3$) capacity. Sludge application rate assumed 7 dry tons (6.4 Mg) per hour, including loading time. The land application vehicle was depreciated on a straight-line basis over a 7-year period. Sludge hauling was based on current rental costs for equipment. Dewatered sludge was assumed to be 22 percent dry solids.

Table 1–37.—Capital cost for vacuum filtration facilities for a 40 Mgal/d (1.75 $m^3/s$) wastewater treatment plant

| | |
|---|---|
| Site work, earthwork, yard piping | $297,000 |
| Chemical storage and feed | 177,000 |
| Filtration equipment | 546,000 |
| Filter and chemical building | 230,000 |
| Sludge loading pad | 78,000 |
| Electrical and instrumentation | 322,000 |
| Subtotal construction cost | 1,650,000 |
| Engineering | 140,000 |
| Total capital cost | 1,790,000 |
| Amortized cost at 30 years., 7 percent int. (CFR = 0.081) | 145,000 |
| Annual capital cost per unit feed dry solids | 8.63 |

Table 1–38.—*Vacuum filtration capital and annual operation and maintenance costs for a 40 Mgal/d (1.75 $m^3/s$) plant*

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Variable cost | | |
| Electric power | $7,100 | $0.42 |
| Chemicals | | |
| Lime | 91,400 | 5.44 |
| $FeCl_3$ | 52,000 | 3.09 |
| Maintenance materials | 7,800 | 0.46 |
| Maintenance labor | 25,800 | 1.54 |
| Laboratory | 6,000 | 0.36 |
| Subtotal variable cost | 190,100 | 11.31 |
| Operator labor | 47,000 | 2.80 |
| Supervision | 15,000 | 0.89 |
| Capital | 145,000 | 8.63 |
| Subtotal fixed cost | 207,000 | 12.32 |
| Total annual cost | 397,100 | 23.63 |

Anaerobically digested sludge land requirements were less than for lime stabilized sludge because of the volatile solids reduction during digestion. The fertilizer value and land rental return credits were taken as previously described in the 4 Mgal/d (0.18 $m^3/s$) design case. Table 1–39 summarizes the total land application cost.

To summarize, the total cost for the lime stabilization and anaerobic digestion alternatives, including ultimate disposal, is shown in table 1–40.

In the 4 Mgal/d case (0.18 $m^3/s$), the total annual cost for the lime stabilization alternative is $74.16 per dry ton ($81.75/Mg) compared to $86.77 per dry ton ($95.65 Mg) for anaerobic digestion. Each of these alternatives assumed liquid application to farmland, with a 3–5 mile (5–8 km) round trip hauling distance. With increasing haul distances, lime stabilization will be decreasingly cost effective because of the greater volume of sludge which must be transported.

In the 40 Mgal/d case (1.75 $m^3/s$), the total annual cost for lime stabilization alternative is $62.94 per dry

Table 1–39.—Annual cost for land application of dewatered anaerobically digested sludges for a 40 Mgal/d (1.75 $m^3/s$) plant

| Item | Total annual cost | Annual cost per ton dry solids |
|---|---|---|
| Amortized cost of land | $202,500 | $12.05 |
| Truck depreciation (spreader only) | 12,100 | 0.72 |
| Truck drivers | 67,100 | 3.99 |
| Truck and loader operation | 260,600 | 15.51 |
| Laboratory | 4,500 | 0.27 |
| Fertilizer credit | (83,400) | (4.96) |
| Land credit | (57,000) | (3.39) |
| Total annual cost | 406,400 | 24.19 |

ton ($69.38/Mg) compared to $94.56 per dry ton ($104.23/Mg) for anaerobic digestion. The cost of pipeline transportation/land application of the liquid sludge is $42.44 per dry ton ($46.78/Mg) compared to $47.82 per dry ton ($52.71/Mg) for dewatering and land application. The pipeline alternative also has the disadvantage of being inflexible for long-term implementation. With the dewatered sludge and truck hauling system, sites could be changed with little difficulty.

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Table 1–40.—Comparison of total annual capital and annual O. & M. cost for lime stabilization and anaerobic digestion including land disposal for a 40 Mgal/d (1.75 $m^3/s$) plant

| | Lime stabilization | | Anaerobic digestion | |
|---|---|---|---|---|
| | Total annual O. & M. cost | Annual cost per ton dry solids | Total annual O. & M. cost | Annual cost per ton dry solids |
| Facilities | | | | |
| Amortized capital lime stabilization | $87,200 | $5.19 | N/A | N/A |
| Amortized capital digesters | N/A | N/A | [a]$756,000 | $45.02 |
| Amortized capital filtration | N/A | N/A | [a]145,000 | [a]8.63 |
| Operating labor | 114,000 | 6.78 | [a]100,000 | [a]5.95 |
| Maintenance labor and materials | 18,300 | 1.09 | [a]90,600 | [a]5.39 |
| Chemicals | 118,600 | 7.06 | [a]143,400 | [a]8.53 |
| Laboratory | 4,500 | 0.27 | [a]12,000 | [a]0.71 |
| Fuel credit | N/A | N/A | [a](71,900) | [a](4.28) |
| Power | 2,000 | 0.12 | [a]7,100 | [a]0.42 |
| Subtotal facilities | 344,600 | 20.51 | [a]1,182,700 | [a]70.37 |
| Land Application | | | | |
| Amortized cost of land, facilities and easements | 339,000 | 20.17 | [a]202,500 | [a]12.05 |
| Truck depreciation | 60,000 | 3.57 | [a]12,100 | [a]0.72 |
| Truck drivers | 201,200 | 11.97 | [a]67,100 | [a]3.99 |
| Truck operations | 263,200 | 15.66 | [a]260,600 | [a]15.51 |
| Pipeline O. & M. | 17,000 | 1.01 | N/A | N/A |
| Power | 35,000 | 2.08 | N/A | N/A |
| Fertilizer credit | (122,700) | (7.30) | [a](83,400) | [a](4.96) |
| Land credit | (84,000) | (5.00) | [a](57,000) | [a](3.39) |
| Laboratory | 4,500 | 0.27 | [a]4,500 | [a]0.27 |
| Subtotal land application | 713,200 | 42.43 | [a]406,400 | [a]24.19 |
| Total annual cost facilities and land application | 1,057,800 | 62.94 | [a]1,589,100 | [a]94.56 |

[a]Includes cost for digestion and vacuum filtration.

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Chapter 2

# Anaerobic Digestion and Design of Municipal Wastewater Sludges

The claimed advantages of the anaerobic digestion process are:[1,2]

- Low sludge production.
- The production of a useful gas of moderate caloric value.
- A high kill rate of pathogenic organisms.
- Production of a solids residue suitable for use as a soil conditioner.
- Low operating cost.

Table 2–1 indicates the kinds of sludges which have been studied on a full-scale basis.

In the past 50 years municipal wastewater sludge has changed from simple primary sludge of purely domestic origin to complex sludge mixtures (primary, secondary, chemical) of domestic and industrial origins.

At first when design engineers only had to consider a primary sludge, the developed rules of thumb[22] were adequate. As the sludge generated became more complex, more and more systems failed and the process developed a "bad reputation." The use of steady state models in the 1960's,[23–25] dynamic models in the 1970's,[26–31] and research into the basic biochemical processes[32–35] has led to significant improvements both in the design and operation of the process. Still the transfer of data from the laboratory to the real world can be difficult.

Table 2–1.—Type and reference of full-scale studies on high rate anaerobic digestion of municipal wastewater sludge

| | Reference on mesophilic | Reference on thermophilic |
|---|---|---|
| Primary and lime | 3,4 | |
| Primary and ferric chloride | 5 | |
| Primary and alum | 6 | |
| Primary and trickling filter | 7,8 | |
| Primary, trickling filter and alum | 9 | |
| Primary and waste activated | 10,11,12 | 11,13,14 |
| Primary, waste activated and lime | 15,16 | |
| Primary, waste activated and alum | 15,17,18 | |
| Primary, waste activated and ferric chloride | 15 | |
| Primary, waste activated and sodium aluminate | 17,18 | |
| Waste activated only (pilot plant only) | 19,20,21 | 19,20,21 |

## GENERAL PROCESS DESCRIPTION

Anaerobic digestion of municipal wastewater sludge is a two-step, very complex biochemical process, dependent on many physical (temperature, solids concentration, degree of mixing, organic loading, detention time) and chemical (pH, alkalinity, volatile acid level, nutrients, toxic materials) factors. Probably the easiest way to visualize what is taking place is to think in terms of a two-step process.

In the first step, facultative microorganisms (sometimes called acid forming bacteria) convert complex organic waste sludge substrate (proteins, carbohydrates, lipids) into simple organic fatty acids by hydrolysis and fermentation. The principal end products, with sludge as substrate, are acetic acid, approximately 70 percent, and propionic acid, about 15 percent.[36–38] The microorganisms involved can function over a wide environmental range and have doubling times normally measured in hours.

In the second step, strictly anaerobic microorganisms (sometimes called methane-forming bacteria) convert the organic acids to methane, carbon dioxide and other trace gases. The bacteria involved are much more sensitive to environmental factors than step one bacteria and normally have doubling times measured in days. Because of this, step two bacteria control the overall process.

Figure 2–1 gives an overview of the entire process. For a more complete review the reader is referred to either Kirsch[35] or Toerien.[32]

## MESOPHILIC—THERMOPHILIC DIGESTION

Temperature can be considered one of the most important factors in the anaerobic digestion process. Even though the total temperature range for operation of the process is very broad, specific microorganisms often have relatively narrow temperature ranges in which they can grow.

For the purpose of classification, the following three temperature zones of bacterial action will be used throughout this chapter:

| | |
|---|---|
| Cryophilic zone | Liquid temperature below 10°C (50°F) |
| Mesophilic zone | Liquid temperature between 10°C to 42°C (50°F to 108°F) |
| Thermophilic zone | Liquid temperature above 42°C (108°F) |

| Raw sludge + | Micro-organisms "A" | $\xrightarrow{K_1}$ Non-reactive products + | Reactive products + | Micro-organisms "B" | $\xrightarrow{K_2}$ $CH_4$ + $CO_2$ | Other end products |
|---|---|---|---|---|---|---|
| Complex substrate<br>Carbohydrates, fats and proteins | Principally acid formers | $CO_2$, $H_2O$<br>Stable and intermediate degradation products<br>Cells | Organic acids<br>Cellular and other intermediate degradation products | Methane fermenters | | $H_2O$, $H_2S$<br>Cells and stable degradation products |

Figure 2–1.—Summary of anaerobic digestion process.[39]

In the past, the vast majority of lab, pilot, and full-scale research has been done in the mesophilic range and some in the thermophilic range. The reason for this is that thermophilic digestion did not seem economical because of the higher energy requirements and the general feeling that operation at the higher temperature would be highly unstable. Recently, the literature indicates that there is a renewed interest in thermophilic digestion[40] because of its elimination of pathogens, high reaction rates and possibly higher gas yields and better dewaterability.

## VOLATILE SOLIDS REDUCTION

One of the main objectives of the anaerobic digestion process is to reduce the amount of solids that need to be disposed. This reduction is normally assumed to take place only in the volatile content of the sludge and it is probably safe to assume only in the biodegradable volatile fraction of the sludge. Research into the area of the biodegradable fraction is quite limited but the following generalities can be used:

1. Approximately 20 to 30 percent of the influent suspended solids of a typical domestic wastewater is nonvolatile.[41] Of the remaining suspended solids that are volatile, approximately 40 percent are nonbiodegradable organics consisting chiefly of lignins, tannins, and other large complex molecules.
2. For waste activated sludges generated from systems having primary treatment, approximately 20 to 35 percent of the volatile solids produced are nonbiodegradable.[42,43]
3. For waste activated sludges generated from the contact-stabilization process (no primaries—all influent flow into aeration tank), 25 to 35 percent of the volatile suspended solids are *nonbiodegradable.*[44]

Though it is realized that only the biodegradable fraction can actually be destroyed, all past research and most of the present day work report on volatile solids destroyed without making any distinction between biodegradable and nonbiodegradable. Because of lack of data, all reference here to solids destruction will be based on volatile solids only.

Figures 2–2, 2–3, and 2–4 show the effect of sludge age and temperature on volatile solids reduction for three common sludges.

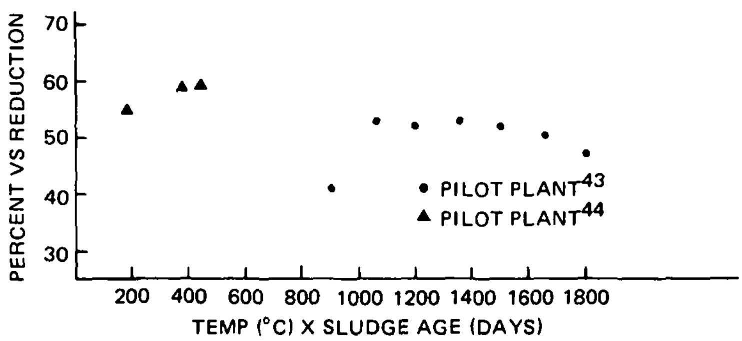



Figure 2–2.—Volatile solids versus reduction versus temperature × sludge age for anaerobically digested primary sludge.

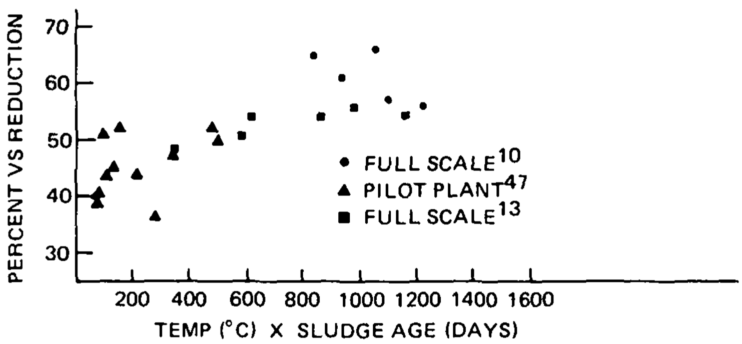



Figure 2–3.—Volatile solids versus reduction versus temperature × sludge age for anaerobically digested mixture of primary and waste-activated sludge.

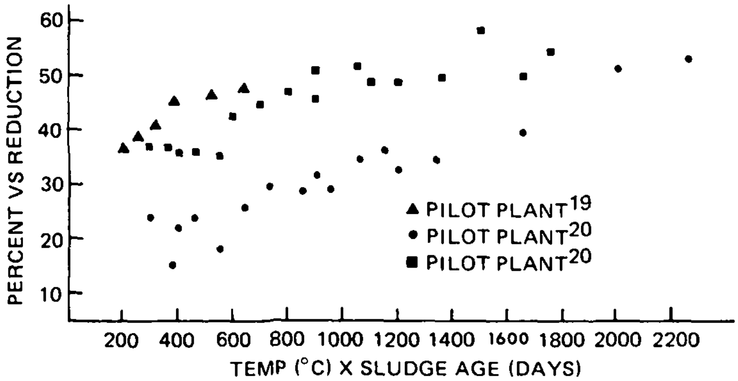



Figure 2–4.—Volatile solids versus reduction versus temperature × sludge age for anaerobically digested waste-activated sludge.

Though the data is somewhat scattered, the following generalizations seem valid.

1. For all three sludges the practical upper limit of volatile solids reduction seems to be 55 percent. Though it was noted that approximately 60 percent of the volatile solids are biodegradable, figures 2–2, 2–3, and 2–4 suggest that practically all the biodegradable fraction is being consumed.
2. The data seem to indicate that under the same design conditions primary sludge will degrade faster than a mixture of primary and waste activated, which in turn degrade faster than straight waste activated. This implies that adjustments must be made in design depending on the type of sludge to be processed.

## SOLIDS CONCENTRATION—ORGANIC LOADING—SLUDGE AGE

Considerable capital cost savings could be realized if the anaerobic digestion process could be operated at higher organic loadings and shorter detention times than commonly used today.

There have been several pilot plant studies[48–51] which have been able to operate at levels approaching 4–5 days residence times, organic loadings approaching 0.5 lb volatile solids/cu ft/day (8.0 kg/m$^3$/day) and solids concentrations up to 12–15 percent solids. Unfortunately, pilot plant digesters are ideally mixed and environmentally controlled, and scaling up the results can be difficult.

Nevertheless, over the years there have been several full-scale facilities which were and still are being operated successfully at short detention times, high organic loadings and high solids concentrations. Some of these plants are listed in table 2–2.

*Solids concentration.*—It must be remembered that the solids concentration within the digester affects the viscosity which in turn affects the ability of the mixing equipment (see section on mixing). Also, because of the solids reduction taking place, the solids concentration within the digester is less than the feed solids concentration. Though it depends on sludge type, the practicable upper limit on the feed solids concentration is in the range of 8 to 9 percent. With a properly designed mixing system this will not cause any operational problems within the digester.

Table 2–2.—Concentration—organic loading—time parameters for several full-scale anaerobic digestion facilities

| Feed solids concentration, percent | Organic load lbs VS/ft$^3$/day | Hydraulic retention time[a] days | Reference |
|---|---|---|---|
| 6.0 | 0.16 | 15.0 | 10 |
| 6.6 | 0.17 | 14.4 | 52 |
| 6.9 | 0.15–0.38 | 11.7–15.9 | 53 |
| | 0.28 | 8.0 | 54 |
| 4.6–5.2 | 0.13–0.17 | 14 | 12 |
| 5.0 | 0.3 | 10 | 12 |
| 6.3 | 0.16 | 16.5 | 55 |
| 8.0 | 0.15 | 21.0 | 56 |

[a] All data based on primary digester only. Digester equipped with mixing and sludge heating.

*Organic loading rate.*—The organic loading rate is a function of the solids concentration within the digester and system sludge age. These two parameters are implicit when one speaks of a loading rate of pounds volatile solids per cubic foot per day. As is shown in table 2–2 designing a digestion system to operate at 0.15 to 0.20 lb VS/cu ft/day (2.4 to 3.2 kg/m$^3$/day) is no problem.

*Sludge age.*—At present, high-rate (mixed and heated) primary digesters to recycle concentrated digested solids are not practiced; therefore, hydraulic residence time and sludge age are almost synonymous. As noted in table 2–2, a minimum time of 15 days in the primary digester is very practicable. It should be remembered though that this time is also related to sludge type and tank temperature, as was shown in figures 2–2, 2–3, and 2–4.

There seems to be an important relationship between the above design parameters. In a study conducted by Clark,[51] involving solids concentration, organic loading rate, and sludge age, the curve shown in figure 2–5 was developed.

The shape of the probable digestion limit curve, i.e., higher organic loadings as the sludge age decreases, is a reflection of the accumulation of various system by-products which may reach inhibitory concentration levels. If for a given digester volume and organic loading rate the sludge age is increased (only possible if influent

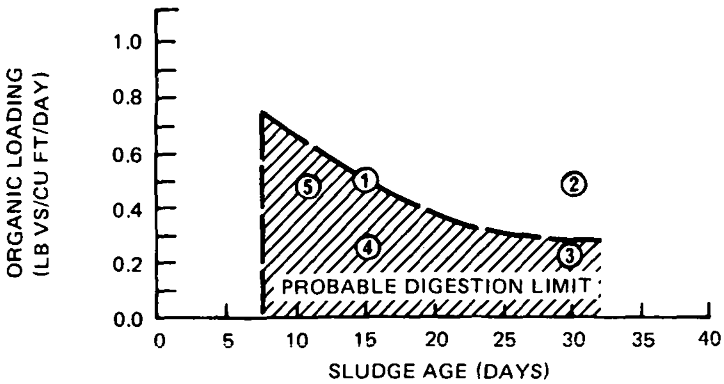



Figure 2–5.—Relationship between solids concentration—organic loading—sludge age limits for anaerobic digestion.[51]

sludge concentration is increased), then the chance for potential inhibitory byproduct concentration levels also is increased.

An engineer designing a high-rate primary digester might, e.g., determine that an organic loading rate of 0.5 lb VS/cu ft/day (8.0 kg/$m^3$/day) and a sludge age (hydraulic detention time) of 15 days is possible (figure 2–5, point 1). If the detention time were doubled (point 2) by doubling the influent solids concentration (volume of tank stays the same), the digester would fail. If instead, the tank volume is doubled (point 3) rather than doubling the influent solids concentration, the unit would still be operating on the failure boundary and nothing would be gained. As a third alternative, halving the loading rate by doubling the tank volume (point 4), assuming the influent solids concentration is halved, would be acceptable. Finally, if the loading rate is to be maintained at 0.5 lb VS/cu ft/day (8.0 kg/$m^3$/day), the sludge age (hydraulic detention time) should be decreased since the tank volume is fixed to allow a lower influent solids concentration.

## MIXING

Mixing in an anaerobic digester that treats municipal wastewater sludge of domestic origin is considered to have the following benefits. (Note: It is assumed that a favorable environment exists to allow development of an anaerobic digestion system.)

- It keeps the food supply uniformly dispersed and in constant contact with the growing cells to promote maximum utilization of the system.
- It keeps the concentration of biological end products at their lowest value by dispersing them uniformly throughout the digester.
- It provides environmental uniformity (temperature, nutrients, etc.) throughout the digester allowing best possible cell development.
- It allows fairly fast dispersion of any toxic material entering the system, thus, possibly minimizing its effect on the anaerobic process.
- It assists in the prevention of a scum layer buildup at the top of the digestion tank.

At the present time not many in the environmental engineering field would dispute the advantages of mixing in an anaerobic digester; however, problems arise with such questions as what is adequate mixing, how do you define mixing, how do you specify mixing, etc.

Before any discussion about mixing can be developed, some time must be spent discussing what and where this mixing is to take place.

### Defining Mixing

In recent years it has become popular to use the term "complete mix" when discussing biological process reactors. Unfortunately, engineers associate this term on a time scale as applied to activated sludge systems when talking about mixing an anaerobic digester.

The term "complete mix" means that the time for dispersion of the feed stream is short in relation to the total hydraulic residence time in the reactor. It is also defined as sufficient mixing so that concentration gradients of chemical and biological ingredients are uniform for the particular reaction rates that exist in the basin.

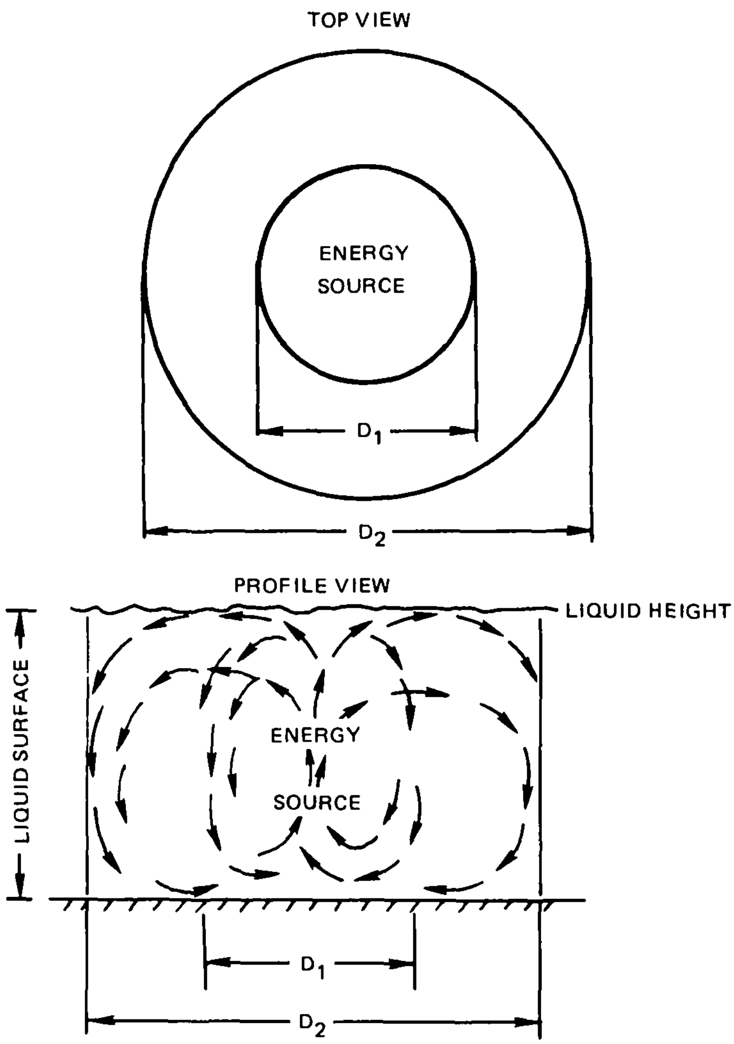
Present-day "complete mix" activated sludge systems have hydraulic residence times of approximately 3 hours based on plant influent flow. Generally a "turn over rate" of 15 to 20 minutes is considered sufficient to achieve "complete mix" conditions within the aeration basin. This would give a turnover rate to hydraulic detention time ratio of 0.08. Present-day high-rate primary digesters have hydraulic detention times of 12 to 17 days. This would seem to imply that a "turnover rate" of about 1 day would provide complete mix conditions within the system.

Mixing within the anaerobic digestion tank occurs on two levels: macromixing and micromixing.[35] Macromixing deals with the bulk mass flow within the digester, while micromixing deals with the degree of intermingling of the system molecules. In biological theory, "complete mix" assumes micromixing.

The actual mixing of the sludge within the digester can be by gas recirculation, mechanical, or a combination of the two. Malina and Miholites[60] describe all present-day systems.

No matter what type of device is utilized the intent is to achieve mixing through a pumping action. Because of this relationship, engineers have come to use the term *horsepower/unit volume* as some type of parameter to define mixing in an anaerobic digester. Unfortunately, this term by itself has no meaning. For mechanical type mixers the wide variation in impeller diameters and speeds can result in similar horsepower but widely different pumping capacities. For gas mixing systems, gas flow, depth, and bubble size can also result in similar horsepower but widely different pumping capacities.

Probably the best way to evaluate mixing is from the standpoint of *zone influence* (figure 2–6): energy is dissipated with movement horizontally away from the energy source. The loss due to friction between the fluid mole-




$D_1$ = EFFECTIVE ZONE DIAMETER FOR MICROMIXING.

$D_2$ = EFFECTIVE ZONE DIAMETER FOR MACROMIXING.

Figure 2–6.—Shear-stress relationship for a thixotropic pseudo plastic material.

cules is a function of liquid density, temperature, and solids concentration. Within a certain area of the point source there is sufficient energy to achieve micromixing. There is also a larger area where bulk flow (macromixing) still takes place even though there is insufficient energy for micromixing.

Presently, the only published work that could be found discussing this concept in the sanitary engineering field was done by the EPA.[61,62] This concept indicates that the width of the micromixing zone in water is no more than twice the liquid depth, with the liquid depth being a function of the type of mixing device utilized and not necessarily the tank liquid depth. It is probably safe to assume that for thickened, anaerobically digested sludges, the zone of influence for any given energy input is smaller than for mixing plain water.

## CHARACTERISTICS OF AN ANAEROBIC DIGESTER

The existing trend in wastewater treatment is to remove more and more material from the main liquid processing stream. This is done through the use of secondary biological treatment schemes, chemical addition, and filters. The sludge produced can vary widely and change rapidly even on an hour-to-hour basis.

Table 2–3 gives specific gravity and particle size distribution on two common type sludges: plain primary and plain waste activated.

There is little data on the rheology of municipal wastewater sludge and even less on anaerobically digested sludge.[48,59] One of the main problems is that it is extremely difficult to do such studies correctly.[58]

Even though the majority of raw wastewater sludges behaves as a thixotropic (time dependent), pseudo plastic, material (figure 2–7), it may not be correct to assume that the sludge within the anaerobic digester has

Table 2–3.—General characteristics of raw primary and waste activated sludge[57]

| | Primary sludge | Waste activated sludge |
|---|---|---|
| Specific gravity | 1.33–1.4 | 1.01–1.05 |
| Particle size | 20% < 1 µm | 40% 1–50 µm |
| | 35% 1–100 µm | 60% 50–180 µm |
| | 45% > 100 µm | |
| Physical appearance | Fibrous | Slimy, gelatinous |

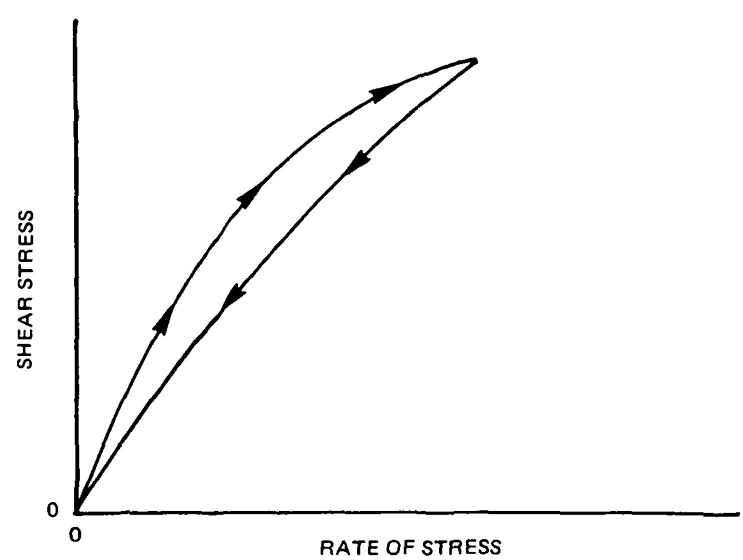



Figure 2–7.—Schematic of zone of mixing influence for energy source in fluid with only fixed upper and lower boundaries.

the same general properties. The liquid within the tank is normally at a higher temperature where there is entrapment of gas bubbles and a general reduction in particle size, all of which affect fluid viscosity.

At the present time anaerobic digestion tanks have a ratio of inside tank diameter to average liquid depth of 1:1 to 5:1. This imposes some restrictions on the ability to develop a mixing regime.

## SUPERNATANT

Poor quality anaerobic digester supernatant is a major operational problem at many municipal wastewater treatment plants. The supernatant will most likely contain high concentrations of carbonaceous organic materials, dissolved and suspended solids, nitrogen, phosphorus, and other materials[63] to impose extra loads on other treatment processes and effluent receiving waters. An analysis of 20 high-rate, mesophilic, two-stage, anaerobic digestion systems[63] showed a range of supernatant suspended solids from 100 to 32,400 mg/l with an average value of 5500 mg/l and for $BOD_5$ from 100 to 6,000 mg/l with an average value of 875 mg/l. Table 2–4 indicates the effects at one midwestern treatment facility where anaerobic digester supernatant from a high-rate system was returned to the plant influent.

Many supernatant treatment alternatives have been tried,[65] some working with a certain degree of success. The question that really needs to be asked is why even expect a clean supernatant stream when digesting mixed primary and waste activated sludges.

The concept of obtaining high quality supernatant developed during the early days of separate anaerobic digestion systems. During this time period the only sludge being digested was primary sludge, which had excellent settling properties (table 2–3).

Modern-day sludges are much more complex. They contain not only primary sludge but sludges generated from secondary treatment, predominately activated sludge systems. Waste activated sludge tends to have fragile floc and is difficult to concentrate by gravity thickening. Because of this, waste activated sludges are thickened by dissolved air flotation thickeners.

Also present-day, high-rate anaerobic digesters are mixed. This constant mixing of the sludge tends to reduce particle size. At the same time the process itself is reducing particle size through biological destruction.

Finally anaerobic digestion systems generate gas throughout the entire tank under a slightly positive pressure (6- to 15-inches (15–38 cm) water column). Thus, the system becomes supersaturated with digester gas.

When the digested sludge is finally pumped to the secondary digester, it contains many fines or sludge that is difficult to gravity thicken and is supersaturated with gas. The gas is then liberated in the form of small gas bubbles which tend to attach themselves to the sludge particles, thus promoting a flotation effect. The combination of these events is very detrimental to gravity concentration. It has been estimated that at least 30 or more days[12] would be required in a secondary digester to obtain a clear supernatant from high-rate systems digesting sludges containing waste activated sludge.

In many cases, it would be better to take all digester contents directly to mechanical dewatering and eliminate provisions for gravity solids-liquid separation. This would give a constant, predictable centrate stream having low suspended solids content.

## ENERGY

### Energy Production

One of the advantages of anaerobic digestion of municipal wastewater sludge is that energy is produced rather than consumed and could go a long way in meeting energy requirements at wastewater plants.[66] One problem encountered with this energy source is in predicting how much energy will be produced for any given plant. This variability in possible production is indicated in table 2–5.

Figure 2–8 shows temperature effects on anaerobic digestion. Schwerin[71] reviewed the literature and plotted

Table 2–4.—Effect of returning supernatant from high rate anaerobic digester to plant influent[64]

| | To primaries lb/day | To secondaries lb/day | Final effluent lb/day | Primary sludge lb/day | Waste activated sludge lb/day |
|---|---|---|---|---|---|
| Suspended solids | 15,969 | 9,501 | 2,836 | 13,249 | 9,593 |
| | [a](36,801) | (15,306) | (3,467) | (19,626) | (14,645) |
| *Total phosphorus* | 914 | 803 | 500 | 156 | 287 |
| | (1,304) | (991) | (435) | (299) | (453) |

[a]Data in parentheses were obtained when untreated anaerobic digester supernatant was discharged to head of plant. Data not in parenthesis were obtained when no supernatant was discharged to head of plant. Data shown is average for the entire time period of study.

Table 2-5.—Cubic feet digester gas produced per pound of organic matter destroyed

| Material | Percent $CH_4$ | $Ft^3$ gas/lb digested |
|---|---|---|
| | Pure compounds[67] | |
| Fats | 62–72 | 18–23 |
| Scum | 70–75 | 14–16 |
| Grease | 68 | 17 |
| Crude fibers | 45–50 | 13 |
| Protein | 73 | 12 |
| | Pure compounds[39] | |
| Carbohydrate | | 14.2 |
| Fat | | 24.6 |
| Insoluble soap | | 22.3 |
| Protein | | 9.4 |

Municipal sludges[68]

"The volume of gas produced per pound of volatile solids digested is reported as 17 to 18 cu. ft./lb at the larger and better instrumented plants. Smaller plants report lesser values, sometimes as low as 6 cu. ft./lb. volatile solids destroyed, but these values are probably due to poor measurement techniques."

Municipal sludges[69]

"...maximum gas production of approximately 11 to 12 cu. ft. of gas per pound of total solids destroyed."

Municipal sludges[70]

"In terms of solids digested, the average yield...is about 15 cu. ft. of gas per pound of volatile solids destroyed."

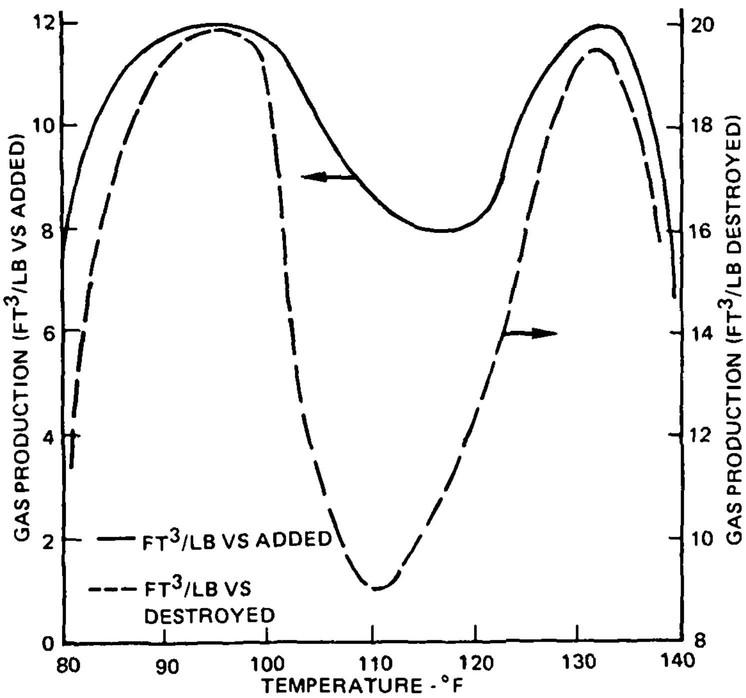



Figure 2-8.—Effect of digestion temperature on gas production based on data from 23 studies.[71]

reported gas production values as a function of digestion temperature. The results show the potential effect of digestion temperature on gas production.

Since the basis of all cost analysis depends on the value of gas produced per mass of solids destroyed, and since there is no existing data, it is suggested that a range of 12 to 17 cu ft/lb (0.75–1.06 $m^3/kg$) volatile solids destroyed be used.

Note: As was noted in the section on Volatile Solids Reduction, the amount of solids destroyed is a function of sludge type and solids retention time (figures 2–1, 2–2, and 2–3).

The heating value of the gas can also range from 550 to 650 Btu/cu ft (20,500–24,200 $kJ/m^3$). Based on an average of 50 plants[72] a value of 600 is suggested.

## Hazards of Digester Gas

*Explosion.*—Sludge gas becomes violently explosive in mixtures of 1 volume gas to 5–15 volumes air. There are many case histories which have shown just how violent and explosive it can be.

*Burning.*—When the ratio of gas to air is higher than the above values, a "burning mixture is encountered." Such a mixture is not as dangerous as an explosive mixture, since it can be extinguished if encountered. However, sewage plant workers have been seriously burned by an instantaneous flame "puff."

*Toxicity.*—Of the minor constituents of sewage gas, hydrogen sulfide ($H_2S$) is the most important. Table 2–6 shows the effects at various concentrations.

*Suffocation.*—Man works best and breathes easiest when the air contains about 21 percent oxygen. Men breathing air that has as little as 15 percent of oxygen usually become dizzy, have a rapid heart beat, and suffer from headache.

Though over 30 years old, two publications by Langford[73,74] on gas safety design considerations are still recommended reading for design engineers. Figure 2–9 shows a schematic of a modern-day gas piping system.[75]

## Digester Gas Utilization

Since digester gas was first used in the United States in 1915[76] for heating and cooking, the use of digester gas has increased, decreased in the 1950's and 1960's because of cheap power alternatives, and presently increasing again because of the energy situation.[77] Several recent publications have described not only operating experience with conventional utilization methods, power generation, and heating[72,77–79] but also potential new

Table 2–6.—Effects of various concentrations of $H_2S$

| | |
|---|---|
| Immediate death | Greater than 2,000 ppm |
| Fatal in 30 minutes or less | 600 to 1,000 ppm |
| Severe illness caused 1/2 to 1 hour | 500 to 700 ppm |
| No severe effects if exposed 1/2 to 1 hour | 50 to 100 ppm |

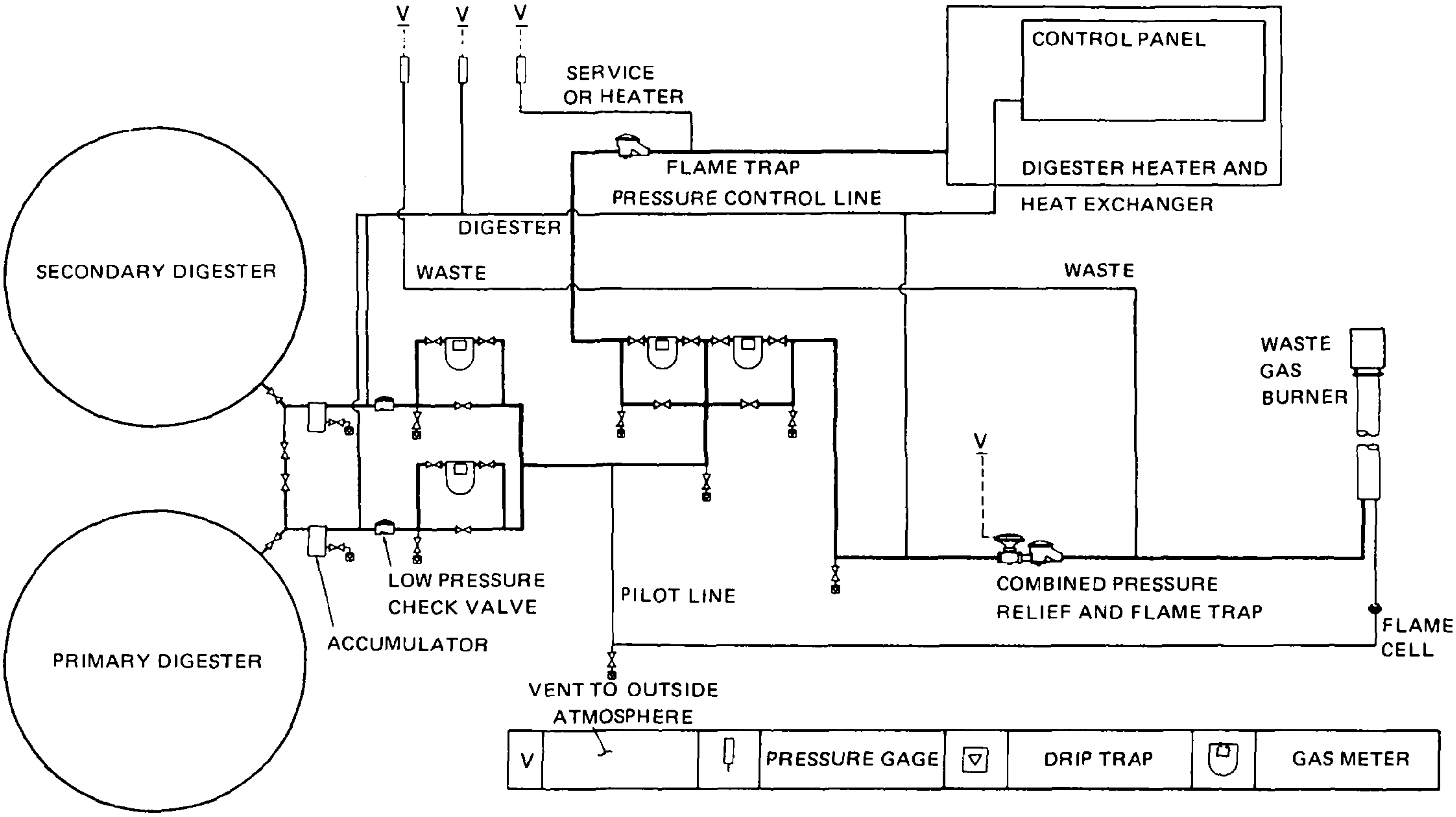



Figure 2–9.—Gas piping schematic of a modern anaerobic digestion system.[75]

uses.[77] One piece of important operating information which has come to light is the amount of hydrogen sulfide permissible for operation of engine generators.[72,79]

Because of its potential corrosive action early uses of digester gas as engine fuel tried to keep $H_2S$ levels under 60 grains per 100 $ft^3$ (200 grains/$m^3$).[80,81] This was done by incorporating some type of dry gas scrubber or wet type bubbling scrubber. Recently a new simple method[82] of removal has been developed.

A recent publication[72] describing the operating results of several plants noted that even though levels of 1,000 to 3,000 mg/l of $H_2S$ were in the gas no adverse effects had been seen on the engines.

## Digester Heat Requirements

In calculating digester heat requirements the two parameters of concern are (see figure 2–10):

1. Heat required to raise the temperature of the incoming sludge flow to digester operating temperature.
2. Heat required to maintain the digester operating temperature (radiation heat loss).

*Heat Required for Raw Sludge.*—It is often necessary to raise the temperature of the incoming sludge stream. The amount of heat required is given by equation 1.

$$Q_s = \frac{\text{gal of sludge}}{\text{day}} \times \frac{8.34 \text{ lbs}}{\text{gal}} \times \frac{(T_2 - T_1)}{1} \times \frac{1 \text{ day}}{\text{hrs}} \qquad (1)$$

where:

$Q_s$ = Btu/hr required to raise incoming sludge stream from temperature $T_1$ to $T_2$
$T_1$ = temperature of raw sludge stream
$T_2$ = temperature desired within the digestion tank
hrs = length of time raw sludge is pumped through the heat exchanger.

*Heat Required for Heat Losses.*—Digesters have radiation heat losses which must be controlled to maintain digester operating temperatures within $\pm 1°F$ otherwise the system could go into thermal shock. The amount of heat loss depends on the tank shape, materials of construction, and external temperatures.

The general design equation for heat flow through compound structures is:

$$Q = U \times A \times (T_2 - T_3) \qquad (2)$$

Figure 2–10.—A heater and heat exchanger.

where:

$Q$ = heat loss Btu/hr
$A$ = area of material normal to direction of heat flow in $ft^2$
$T_2$ = temperature desired within the digestion tank
$T_3$ = temperature outside the digestion tank

$$U = \frac{1}{\sum \frac{1}{C_i} + \sum \frac{x_i}{k_i}} \qquad (3)$$

Table 2–7.—"U" factors for various anaerobic digestion tank materials[75]

| Material | U |
|---|---|
| Fixed steel cover (1/4″ plate) | 0.91 |
| Fixed concrete cover (9″ thick) | 0.58 |
| Floating cover (wood composition) | 0.33 |
| Concrete wall (12″ thick) exposed to air | 0.86 |
| Concrete wall (12″ thick), 1″ air space and 4″ brick | 0.27 |
| Concrete wall or floor (12″ thick) exposed to wet earth (10′ thick) | 0.11 |
| Concrete wall or floor (12″ thick) exposed to dry earth (10′ thick) | 0.06 |

where:

$C_i$ = conductance for a certain thickness of material Btu/hr-$ft^2$–°F
$x_j$ = thickness of material—inches
$k_j$ = thermal conductivity of material Btu – (inch)/hr-$ft^2$–°F

Values of $C_i$ and $k_j$ can be found in various handbooks.[82]

Various values of U for different digester covers, wall construction, and floor conditions are given in table 2–7.

## NUTRIENTS

In general, it is commonly assumed that municipal wastewater sludge is not nutrient deficient. It has been extremely difficult to conduct research on optimum nutrient requirements of anaerobic bacteria on sewage sludge.[84] To date, the literature has shown[85] that, like aerobic bacteria, nitrogen and phosphorus are required in the highest amount (12 and 2 percent, respectively, based on the weight of biological solids present in the system). It is suggested that a minumum C:N:P ratio of 100:15:1 be used for design purposes.

Several researchers have also shown that the addition of certain trace materials, iron[86] and sulfur,[84] could be very beneficial to the process.

## pH CONSIDERATIONS

As was noted under General Process Description, anaerobic digestion is a two-step process consisting of an "acid forming" and "methane forming" step. During the first step the production of volatile acid tends to reduce the pH. The reduction is normally countered by destruction of volatile acids by methane bacteria and the subsequent production of bicarbonate.

Past research[87–89] has shown that the optimum pH value for methane producing bacteria is in the range of 6.4 to 7.5 and that these bacteria are very sensitive to pH change. Recent research though[90] now seems to indicate that the pH tolerance of methane producing bacteria is greater than previously expected. The study also indicated that high and low pH values were only bacteriostatic and not bactericidal as previously thought. Because of the importance of this finding to system control, more research is needed in this area to verify these results.

pH is related to several different acid-base chemical equilibria. In the anaerobic digestion process the range of interest is between 6.0 to 8.0, which for all practical purposes makes the carbon dioxide-bicarbonate relationship the most important. In an anaerobic digestion system the amount of carbon dioxide is dependent only on the law of partial pressure. Since soluble carbon dioxide depends primarily on the $CO_2$ gas content and since at any given time the composition of digester gas is relatively fixed, pH is a function of the bicarbonate concentration as shown in figure 2–11.

This relationship is very important from a process control standpoint.[92] Also, it points out the importance of

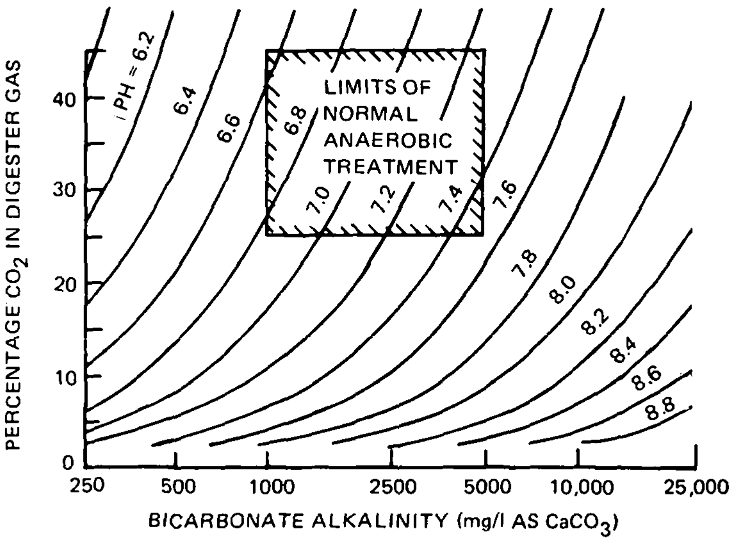



Figure 2–11.—Relationship between pH and bicarbonate concentration near 95°F.[91]

analyzing for bicarbonate alkalinity instead of total alkalinity as is commonly done today. The relationship between the two is given in equation 4.

$$BA = TA - 0.71\ VA \tag{4}$$

where:

- BA = bicarbonate alkalinity as mg/l $CaCO_3$
- TA = total alkalinity as mg/l $CaCO_3$ determined by titration to pH 4.0.
- VA = volatile acids measured as mg/l acetic acid
- 0.71 = a combination of two factors (0.83)(0.85). 0.83 converts volatile acids as acetic to volatile acid alkalinity $CaCO_3$ and 0.85 from the fact that in a titration to pH 4.0, about 85 percent of the acetate has been converted to the acid form.

It has been suggested[92] that the only way to increase digester pH is by the addition of sodium bicarbonate. Other materials such as caustic soda, soda ash, and lime cannot increase bicarbonate alkalinity without reacting with soluble carbon dioxide, which in turn causes a partial vacuum within the system. Also above pH 6.3, lime may react with bicarbonate to form insoluble calcium carbonate, thus promoting scale formation or encrustration.

Sodium can be toxic at certain concentrations (see section on Toxicity—light metal cations), and it is recommenced to keep sodium levels below 0.2M (approximately 4,600 mg/l), which may require dilution of the digester contents as part of the corrective measures.

## TOXICITY

Kugelman and Chin[93] have noted that much of the published data on toxicity in anaerobic digestion systems is erroneous and misleading because of inadequate experimental techniques and general lack of understanding. Therefore, before any discussion of toxicity takes place a review of several fundamentals must be made.

First of all for any material to be biologically toxic it must be in solution. If any substance is not in solution, it is not possible for it to pass through the cell wall and therefore cannot affect the organism.

Second, toxicity is a relative term. There are many organic and inorganic materials which, depending if they meet condition one above, can be either stimulatory or toxic. A good example of this is the effect of ammonia nitrogen on anaerobic digestion—table 2–8.

Acclimation is a third consideration. When potential toxic materials are slowly increased within the environment, many biological organisms can rearrange their metabolic resources, thus overcoming the metabolic block produced by the toxic material. Under shock load conditions there is not sufficient time for this rearrangement to take place.

Finally, there is the possibility of antagonism and synergism. Antagonism is defined as a reduction of the toxic effect of one substance by the presence of another. Synergism is defined as an increase in the toxic effect of one substance by the presence of another. This is an important relationship for cation toxicity.

Though there are many potential toxic materials, this section will only concern itself with the following: volatile acids; heavy metals; light metal cations; oxygen; sulfides; and ammonia.

*Volatile acids.*—Up until the 1960's it was commonly believed that volatile acid concentrations over 2,000 mg/l were toxic to an anaerobic digester. There was also considerable controversy on whether or not alkaline substances should be added to maintain adequate buffer capacity.

In the early 1960's, McCarty and his coworkers published results from very carefully controlled studies in this area.[94,96,97] Their results showed the following:

1. Studies clearly indicate that volatile acids, at least up to 6,000–8,000 mg/l, were not toxic to methane bacteria. Therefore, as long as there was adequate buffer capacity to maintain the system pH in the range of 6.6 to 7.4, the system would function.
2. Control of pH by the addition of an alkaline material is a valid procedure as long as the cation of the

Table 2–8.—Effect of ammonia nitrogen on anaerobic digestion[94,95]

| $NH_3$-N | Effect |
|---|---|
| 50–200 | Beneficial |
| 200–1,000 | No adverse effects |
| 1,500–3,000 | Inhibitory at pH over 7.4–7.6 |
| Above 3,000 | Toxic |

alkaline material does not cause toxicity. It was found that the addition of sodium, potassium, or ammonium compounds is detrimental but magnesium or calcium alkaline compounds are not.

*Heavy metals*—Heavy metal toxicity has frequently been cited as the cause of many anaerobic digestion failures. Even though trace amounts of most heavy metals are necessary for maximum biological development,[98] the concentrations existing in raw wastewater sludges could cause potential problems.

Since heavy metals tend to attach themselves to sludge particles,[99,100] even low influent concentrations can be concentrated significantly in the sludge-handling process. Table 2–9—column 2 gives the range of influent concentrations of some heavy metals. The range is quite wide with the higher values normally being attributed to a local industrial polluter.

Column 3 of table 2–9 gives the typical range of removal that can be expected through a standard secondary treatment system. Published data seem to indicate that the percent removal, without chemical addition, is a function of influent concentration. The higher the influent concentration the higher the percent removal.

Column 4 of table 2–9 shows expected removals with lime additions at a pH of 11.0. In fact it has been noted[106] that treatment systems which add lime or other chemical coagulations for phosphate removal can expect significant amounts of influent heavy metals to also be removed.

Because of the dependence of inhibition on naturally occurring reagents, such as carbonate and sulfide, it is not possible to define precise total toxic concentrations for any heavy metal.[107] Table 2–10 gives some concentrations of individual metals required to cause severe inhibition. Table 2–11 gives an indication of the difference between total and soluble concentrations that may exist in an anaerobic digester.

The problem of heavy metal toxicity may not necessarily be reduced with strict enforcement of industrial point sources. For example, the normal digestion and excretion of zinc is approximately 10 mg per person.[109] Another nonprofit source is the paved street. Table 2–12 gives the results of a study on heavy metal pollution from paved road surfaces of several large cities.[109] In another extensive study,[110] based on 9,600 analyzed samples, it was shown that if all industry in metropolitan New York had zero discharge, there would only be a 9 percent reduction in copper, 20 percent in chromium, 6 percent in zinc, 16 percent in cadmium, and 62 percent in nickel.

Except for chromium, heavy metal toxicity in anaerobic digesters can be prevented or eliminated through precipitation with sulfides.[108,111-113] Hexavalent chromium is normally reduced to trivalent chromium which under normal anaerobic digester pH levels is relatively insoluble and not very toxic.[114]

Table 2–9.—Influent concentrations and expected removals of some heavy metals in wastewater treatment systems

| Heavy metal | Influent concentrations (mg/l) | Removal: Secondary treatment, (percent) | Removal: Lime—pH 11, (percent) |
|---|---|---|---|
| Cadmium | <.008–1.142[101,104] | 20–45[101] | 95[103,109] |
| Chromium +3 | <.020–5.8[101,104] | 40–80[101] | 95[106] |
| Chromium +6 | <.020–5.8[101,104] | 0–10[101] | 20[106] |
| Copper | <.020–9.6[101,104] | 0–70[101] | 90[103,109] |
| Mercury | <.0001–.068[101,104] | 20–75[101] | 40[106] |
| Nickel | <.1–880[101,104] | 15–40[101] | 90[103,106] |
| Lead | <.05–12.2[101,104] | 50–90[101] | — |
| Zinc | <.02–18.00[101,104] | 35–80[101] | 90[103,106] |
| Arsenic | <.002–.0034[102] | 28–73[102] | 70[106] |
| Iron | <.1–13[104] | 72[105] | 99[103] |
| Manganese | <.02–.95[102] | 25[105] | 95[103,106] |
| Silver | <.05–.6[104] | — | 96[103] |
| Cobalt | Below detection[104] | — | — |
| Barium | — | 47[105] | 75[106] |
| Selenium | — | 79[105] | — |

Table 2–10.—Total concentration of individual metals that have been found to cause severe inhibition in anaerobic digesters[107]

| Metal | Concentration of metal in digester contents (dry sludge solids): Percent | mM Kg$^{-1}$ |
|---|---|---|
| Copper | 0.93 | 150 |
| Cadmium | 1.08 | 100 |
| Zinc | 0.97 | 150 |
| Iron | 9.56 | 1,710 |
| Chromium +6 | 2.20 | 420 |
| Chromium +3 | 2.60 | 500 |

Table 2–11.—Total and soluble heavy metal content of digesters[108]

| Metal | Total concentrations (mg/l) | Soluble concentrations (mg/l) |
|---|---|---|
| Chromium +6 | 420 | 3.0 |
| Copper | 196 | 0.7 |
| Nickel | 70 | 1.6 |
| Zinc | 341 | 0.1 |

Table 2-12.—Heavy metal from paved-curb streets[109]

| Metal | Arithmetic mean | Range |
|---|---|---|
| Zinc | [1]0.75 | [1]0.062-2.1 |
| Copper | .21 | .020-.59 |
| Lead | .68 | .03 -1.85 |
| Nickel | .060 | .011-.19 |
| Mercury | .080 | .019-.2 |
| Chromium | .12 | .0033-.45 |

[1]Data given in pounds/mile of paved street.

The reason for using sulfide precipitation is the extreme insolubility of heavy metal sulfides.[115] Approximately 0.5 mg of sulfide is required to precipitate 1.0 mg of heavy metal. If insufficient sulfide is not available from natural sources, then it must be added in the form of sulfate which is reduced to sulfide under anaerobic conditions.

One potential drawback of using the sulfide saturation method is the possible production of hydrogen sulfide gas or sulfuric acid due to excess amounts of sulfide in the digester. Because of this, it is recommended that ferrous sulfate be used as a source of sulfide.[93] Sulfides will be produced from the biological breakdown of sulfate, and the excess will be held out of solution by the iron. However, if heavy metals enter the digester, they will draw the sulfide preferentially from the iron because iron sulfide is the most soluble heavy metal sulfide.

Two other methods of controlling excess sulfide additions have been proposed.[112,116] One method would be to continuously analyze the digester gas for hydrogen sulfide.[103] When there are detectable levels of $H_2S$, sulfate addition would be terminated; when the level becomes undetectable, additions would start. A second method[116] was the use of a silver-silver sulphide electrode to measure very low levels of soluble sulphides. The electrode is calibrated in standardized solutions of sodium sulphide of known value to yield a parameter, pS, defined in a manner similar to pH, as the negative common logarithm of the divalent sulphide ion concentration. For example, when $S^{-2}$ is $10^{-5}$M, pS would be 5.

*Light metal cations*—Only recently[93,117,118] has the significance of the light metal cations (sodium, ammonium, potassium, magnesium, calcium) in anaerobic digestion started to be unraveled. Normally, domestic wastewater sludges have low concentrations of these cations but significant contributions, enough to cause toxicity, can come from two sources.

1. Industrial operations.
2. The addition of alkaline material for pH control.

Not only can each of these cations be either stimulatory or toxic depending on concentration (table 2-13) but when combined with each other will produce either an antagonism or synergism relationship.

Table 2-13.—Stimulatory and inhibitory concentrations of light metal cations[117,118]

| Cation | Stimulatory (mg/l) | Moderately inhibitory (mg/l) | Strongly inhibitory (mg/l) |
|---|---|---|---|
| Calcium | 100-200 | 2,500-4,500 | 8,000 |
| Magnesium | 75-150 | 1,000-1,500 | 3,000 |
| Potassium | 200-400 | 2,500-4,500 | 12,000 |
| Sodium | 100-200 | 3,500-5,500 | 8,000 |

Based on current knowledge whenever inhibition is being caused by an excess of a certain cation, the cation can be antagonized by the addition of one or more of the cations listed in table 2-14.

*Oxygen.*—Engineers have always been concerned with air getting into anaerobic digesters since a mixture of one volume digester gas with 5 to 15 volumes of air is an explosive mixture.

Many engineers have also expressed concern over the possibility of oxygen toxicity when using dissolved air flotation thickeners for sludge thickening. In 1971 Fields and Agardy[119] showed "... that small additions of air (up to 0.01 volume per volume of digester contents) approaching one percent by volume, will not significantly affect anaerobic digester performance." This value is several magnitudes higher than the amount of air that would be generated from a dissolved air thickening system.

*Sulfides.*—By itself soluble sulfide concentrations over 200 mg/l are toxic to anaerobic digestion systems.[111,120] The soluble sulfide concentration within the digester is a function of the incoming source of sulfur, the pH, the rate of gas production, and the amount of heavy metals to act as complexing agents. The high levels of soluble sulfide can be reduced by the addition of iron salts, or gas scrubbing.

*Ammonia.*—Whenever there are high concentrations of protein waste, which is possible in some systems with highly concentrated feed sludges, ammonia toxicity must be considered.[94,118] Ammonia can be in two forms, ammonium iron $NH_4+$ or ammonia gas. Both forms are always

Table 2-14.—Cation antagonists

| Inhibiting cation | Antagonist cation |
|---|---|
| Ammonium | Potassium |
| Calcium | Sodium, potassium |
| Magnesium | Sodium, potassium |
| Potassium | Sodium, potassium, calcium, ammonium |
| Sodium | Potassium |

in equilibrium, the concentration of each depending on pH. Equation 5 shows the relationship.

$$NH_4^+ \rightleftarrows NH_3 + H^+ \quad (5)$$

When the pH is 7.2 or lower, equilibrium is shifted toward the ammonium ion and inhibition is possible at certain concentrations. At pH values over 7.2, the reaction shifts toward the gas phase which is inhibitory at low values.

Analysis for ammonia toxicity is done by analyzing the total ammonia concentration. If the total ammonia concentration is between 1,500 to 3,000 mg/l and the pH is above 7.4–7.6, there are possible inhibitory effects due to ammonia gas. This can be controlled by the addition of enough HCl to maintain the pH between 7.0 to 7.2. If total ammonia levels are over 3,000 mg/l, then the $NH_4+$ion will become toxic no matter what pH level. The only solution is to dilute the incoming waste sludge.

## BACTERIAL EFFECTS

Pathogenic organisms in wastewaters consist of bacteria, virus, protozoa, and parasitic worms. Many of these organisms, especially enteric viruses,[122] have a strong tendency to bind themselves to sludge solids.

Table 2–15.—Human enteric pathogens occurring in wastewater and the diseases associated with the pathogens[125]

| Pathogens | Diseases |
|---|---|
| Vibrio cholera | Cholera |
| Salmonella typhi | Typhoid and other enteric fevers |
| Shigella species | Bacterial dysentery |
| Coliform species | Diarrhea |
| Pseudomonas species | Local infection |
| Infectious hepatitus virus | Heptatitis |
| Poliovirus | Poliomyletis |
| Entamoeba histolytica | Amoebic dysentery |
| Pinworms (eggs) | Asheariasis |
| Tapeworms | Tapeworm infestation |

Table 2–16.—Pathogenic organisms in sludge[123,124]

| Type | Salmonella (No./100 ml) | Pseudomones aeruginosa (No./100 ml) | Fecal coliform (No. $\times 10^6$/100 ml) |
|---|---|---|---|
| Raw primary | 460 | $46 \times 10^3$ | 11.4 |
| | 62 | 195 | |
| Trickling filter | 93 | $110 \times 10^3$ | 11.5 |
| Raw WAS | 74 | $1.1 \times 10^3$ | 2.8 |
| | 2,300 | $24 \times 10^3$ | 2.0 |
| | 6 | $5.5 \times 10^3$ | 26.5 |
| Thickened raw WAS | 9,300 | $2 \times 10^3$ | 20 |

Table 2–17.—Pathogenic organisms in mesophilic anaerobically digested sludge[123,124]

| | Salmonella (No./100 ml) | Pseudomonas aeruginos (No./100 ml) | Fecal coliform s $10^6$ (No./100 ml) |
|---|---|---|---|
| Primary only | 29 | 34 | 0.39 |
| WAS only | 7.3 | $10^3$ | 0.32 |
| Mixture | | | |
| Primary and WAS | 6 | 42 | .26 |

Table 2–15 lists the human enteric pathogens that have been found in wastewater sludges along with the diseases normally associated with them. Table 2–16 lists some data on bacterial concentrations found in raw sludges from two studies.[123,124]

The reduction of pathogenic organisms under mesophilic, anaerobic digestion has been studied by various researchers.[122,126–129] Though some early research indicated die off may be due to bactericidal effects,[126,127] current research supports that die off is strictly related to natural die off. Data from two studies are given in table 2–17 for mesophilic anaerobically digested sludge.

No reported work on pathogen destruction for thermophilic anaerobic digestion could be found.

## ACTIVATED CARBON

The first reported studies on the addition of activated carbon to anaerobic digesters treating municipal wastewater sludges was in 1935, at Plainfield, N.J.,[131] and in 1936 in U.S. Patent 2,059,286.[132] At this time the addition of activated carbon was claimed to have the following benefits:

1. Enhanced the rate of digestion.
2. Increased the total amount of gas produced.
3. Produced clear supernatants.
4. Enhanced the drainability of the digested sludge.
5. Increased temperatures within the digester.
6. Gave higher volatile solids reductions.

Until recently no other reported work in this area could be found. In 1975 Adams[133,134] discussed the results of studies carried out by ICI. In his discussion he pointed out the following advantages based on full-scale studies carried out at Cranston, R.I.[135] and Norristown, Pa.[136]

1. Promoted sludge settling and clear supernatants due to the high carbon density.
2. Catalyzes the breakdown of sludge solids, thereby reducing the amount of sludge to be handled.
3. Increase gas production per mass of solids added plus producing a gas with higher methane content.
4. Can absorb certain substances such as pesticles, heavy metals, grease, scum, and detergents.

5. Reduction in odors.
6. Possible improvement in mechanical dewatering operation at least for vacuum filtration.

Even though several full-scale studies have been conducted, they have not been done scientifically but more of a general "add some carbon and see what happens" attitude. Though improved operating results have been shown, the real mechanism for these results have not yet been clearly identified. At the present time EPA has awarded a grant to Batelle to study the effects of activated carbon addition on anaerobic digesters.

## TANK LAYOUT

Essentially four basic types of anaerobic digestion systems are available to stabilize municipal wastewater sludges. The four systems are discussed below in order of their complexity.

*Conventional low rate anaerobic digestion.*—Figure 2–12 shows what is typically thought of as a conventional, low rate, anaerobic digestion system. Essentially, this system is nothing more than a large storage tank and no attempt to control the environment or accelerate the process is made.

*Conventional high rate anaerobic digestion.*—Figure 2–13 shows what is typically considered a conventional, high rate, anaerobic digestion system and is the most commonly used system in the United States today. In this system attempts are made to control the environment (through thickening, heating and mixing) and accelerate the process. Essentially, all digestion takes place in the first tank. This tank is normally maintained at 95°F (34°C) and mixed with some type of gas mixing system. Hydraulic detention times are normally 15–25 days. The majority of designs also provide a so-called "secondary digester" for solids-liquid separation (dotted line tank in figure 2–13) but this practice is being challenged as not being useful in many applications and that going direct to mechanical dewatering can have several significant advantages.[137]

*Anaerobic contact.*—The advantage of sludge recycle in the anaerobic digestion process has not only been discussed but applied[138–141] in treating high strength waste and has been indicated to be worthwhile in treating waste sludges.[142] Nevertheless, this process alternative is rarely considered in municipal anaerobic sludge digestion.

Figure 2–14 shows a typical schematic of the process. The essential feature of this system is that positive separation through the use of a centrifuge biomass is utilized. Part of this biomass is recycled back to the anaerobic digester where it is mixed with the incoming sludge. This recycling of the sludge thus allows for adequate cell retention to meet kinetic requirements yet significantly reduces hydraulic detention time.

*Phase separation.*—As was noted under the general process section, the anaerobic digestion process consists of two distinct phases. The previous three systems attempted to do this in one reactor. As early as 1958[143] the possible value of actually separating the two processes was discussed. Work in 1968[144] using dialysis separation techniques clearly showed "—that the hydrolysis-acid production sludge is the rate limiting process in anaerobic digestion of sewage sludge. Furthermore, the acid formers in a digester must operate at below optimum conditions in order to maintain a healthy population of methane forming bacteria." During the past several years considerable research has been conducted in this area which was summarized by Ghosh[145] and has also led to a patented process.[146] Figure 2–15 shows a sche-

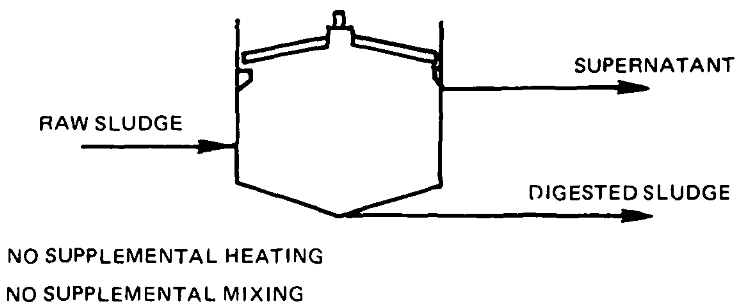



Figure 2–12.—Schematic of conventional low rate anaerobic digestion system.

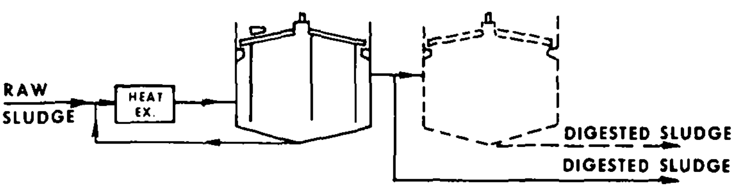



Figure 2–13.—Schematic of conventional high rate anaerobic digestion system.

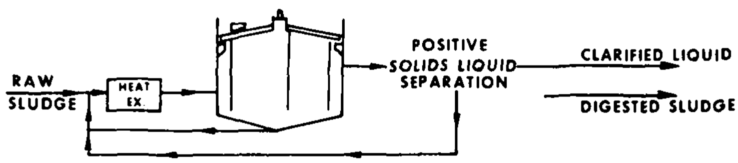



Figure 2–14.—Schematic of anaerobic contact process.

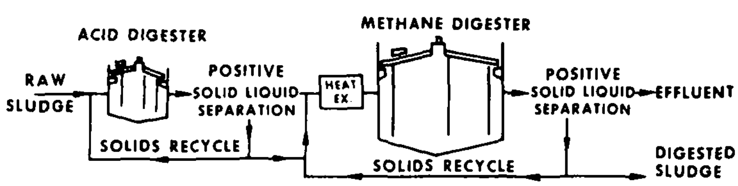



Figure 2–15.—Schematic of phase separation anaerobic digestion of sludge.[145]

matic of this multistage system as conceived by Ghosh.[145]

The phase separation process has several potential benefits when compared to the other processes. These are:[145]

1. Capability of maintaining the optimum environment for each group of digester organisms.
2. Substantial reduction in total reactor volume and the consequent savings in capital and operating costs.
3. Higher rates of solids stabilization and increased production rate and methane content of the final product gases.
4. Decreased heat requirement and increased thermal efficiency.
5. Suitable for incorporation into existing treatment plants with minimum capital investment.
6. Reduction of nitrogen content of the system effluent by simultaneous liquefaction and denitrification of waste feeds in the acid digester.

## GENERAL OPERATIONAL CONTROL PROCEDURES

It should be noted that there is no one test or control parameter that will signify good or bad anaerobic digestion operation. Control or operation of an anaerobic digestion system should be done through a combination of several analyses, the results plotted as a function of time. In this way an unbalanced digester would be defined as one which starts to radically deviate from past norms. Note that the norm at one plant can be failure conditions at another.

At the present time it is suggested that a minimum of four different tests be performed on a regular basis. The four proposed tests are: pH, bicarbonate alkalinity, volatile acids and percent carbon dioxide ($CO_2$) in the digester gas.

*pH.*—As was discussed under pH Considerations, optimal pH is between 6.4 to 7.5. Unfortunately, the pH test by itself is not a good control procedure[92] because:

1. It is a logarithmic function and is not very sensitive to large fluctuations in the alkalinity concentration. For example, a change in alkalinity from 3,600 to 2,200 mg/l would only change the pH from 7.1 to 6.9 which is within the error involved in pH measurement.
2. It does not provide adequate warning. A low pH only informs the operator that an upset has occurred.

*Bicarbonate alkalinity.*—The importance of measuring bicarbonate alkalinity rather than total alkalinity was discussed in the section entitled "pH Considerations." The bicarbonate alkalinity and volatile acid test are used together to develop the ratio of volatile acid to bicarbonate alkalinity. In order to insure good operation (that is good buffering capacity), this ratio should be below 0.7.

Note: A fast, simple method for differentiating bicarbonate and volatile acid alkalinity without using distillation has been developed by DiLallo and Albertson.[147]

*Volatile acids.*—By itself this analysis means nothing. Only when plotted as a function of time or used in conjunction with the volatile acid-bicarbonate ratio can impeding operation problems be interpreted early enough to allow some type of correctional procedures.

*Carbon dioxide content.*—Under good operation the $CO_2$ content in digester gas will be between 35–45 percent. As an unbalance condition starts to occur, there will be an increase in the percentage of $CO_2$ as the methane producers become incapable of functioning.

When the control parameters indicate an unbalance condition, the following steps of action have been recommended:[91]

1. Maintain pH near neutrality
2. Determine cause of unbalance
3. Correct cause of unbalance
4. Provide pH control until treatment returns to normal.

Maintaining the pH near neutrality can be done two ways. The first is to reduce the waste feed. A second way is through the addition of some type neutralizing material (see section on pH Considerations and Toxicity—Light Metal Cations).

Determining the cause of unbalance can be difficult. Some of the easier things to check are hydraulic washout, heat exchanger not capable of providing sufficient heat, mixing system not operating, sudden change in the amount of sludge pumped to the digester and extreme drop in pH. If nothing shows up after the above preliminary analysis, then testing for ammonia, free sulfides, heavy metal and light metal concentrations will have to be made.

Once it has been determined what is causing the problem, corrective measures can be taken to put the digester back on line. Depending on the cause of unbalance, the length of time required to bring a digester back to normal operating condition may take from 2 to 3 days to 4 to 6 months.

## BASIC SIZING CRITERIA—ANAEROBIC DIGESTION SYSTEMS

| | |
|---|---|
| Operating temperature for optimizing gas production | Mesophilic 35–37.8° C (95–100° F) |
| | Thermophilic 54.4–57.2° C (130–135° F) |
| Hydraulic detention time (no recycle) of primary digester to achieve max. volatile solids destruction | |
| primary sludge only | 8.5–10 days at 35° C |
| | 5.5–7 days at 54.4° C |
| primary plus waste activated sludge | 15–17 days at 35° C |
| | 11–13 days at 54.4° C |
| waste activated sludge only | 25–27 days at 35° C |
| | 16–18 days at 54.4° C |

| | |
|---|---|
| Feed solids concentration, degritted sludge | Mixing difficulties start to develop at feed solids concentrations over 8–9 percent. |
| Organic loading rate | Function of hydraulic detention time and feed solids concentration. Many present-day facilities are operating from 0.15 to 0.25 lb VS/ft$^3$/day (2.4–4.0 kg/m$^3$/day). |

Note: When dealing with anaerobic digestion of waste activated sludges or mixtures of, should not assume any solids-liquid separation within secondary digester.

## DESIGN PROBLEM

Two designs, a 4 Mgal/d (0.18 m$^3$/s) and 40 Mgal/d (1.75 m$^3$/s), are evaluated. Influent is typical domestic wastewater of 200 mg/liter biochemical oxygen demand ($BOD_5$) and 200 mg/liter suspended solids (SS) with no heavy industrial contributors. Liquid treatment consists of grit removal, primary treatment, secondary treatment (activated sludge), and chlorination. No chemicals are added to liquid treatment portion.

### Sludge Type and Amount

Every million gallons (3,785 m$^3$) of raw plant influent will generate approximately 1,000 lb (453.6 kg) of primary sludge and 1,000 lb (453.6 kg) of waste-activated sludge.[148] This can be further broken down as in table 2–18.

Based on table 2–18, the total sludge generated for the design examples would be 8,000 lb (3,636 kg) for the 4-Mgal/d (0.18 m$^3$/s) design and 80,000 lb (36,364 kg) for the 40 Mgal/d (1.75 m$^3$/s) design.

### Temperature

Operating temperature in a high-rate digester would be:

- 35°C (95°F) for a 4-Mgal/d (0.18 m$^3$/s) design based on mesophilic conditions
- 54.4°C (130°F) for a 40-Mgal/d (1.75 m$^3$/s) design based on thermophilic conditions

The coldest ambient air temperature for both designs is assumed to be 12.2°C (10°F). The coldest raw sludge temperature for both designs is assumed to be 4.5°C (40°F).

Table 2–18.—Breakdown of inert and volatile suspended solids per mg of plant influent (lbs)

| | Inert Nonvolatile | Inert volatile | Biodegradable volatile |
|---|---|---|---|
| Primary sludge | 250 | 300 | 450 |
| Waste-activated sludge | 300 | 210 | 490 |
| Totals | 550 | 510 | 940 |

### Required Hydraulic Residence Time—Organic Loading—Influent Solids Concentration for High-Rate Digester

For both designs maximum volatile solids destruction is desired. Figure 2–3 shows that for this particular type sludge, a practical upper limit of 55 percent volatile solids destruction is possible and can be obtained in 600 degree-days.

Thickened sludge recycle will not be used in either design; therefore, sludge age will equal hydraulic residence time (HRT) in a high-rate digester.

4 Mgal/d design

600°C − days ÷ 35°C = 17 days minimum HRT.

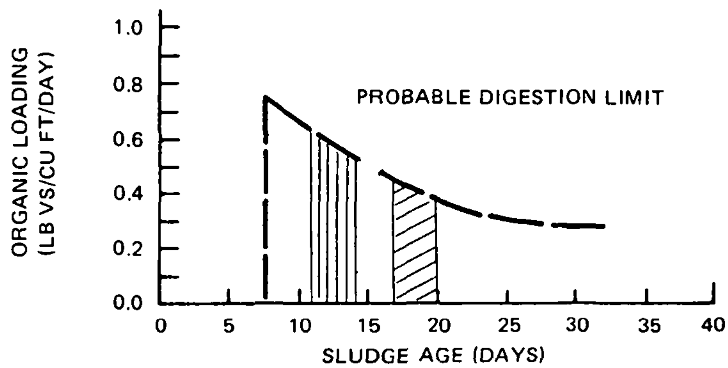



Figure 2–16.—Relationship between solids concentration—organic loading—sludge age for anaerobic digestion.

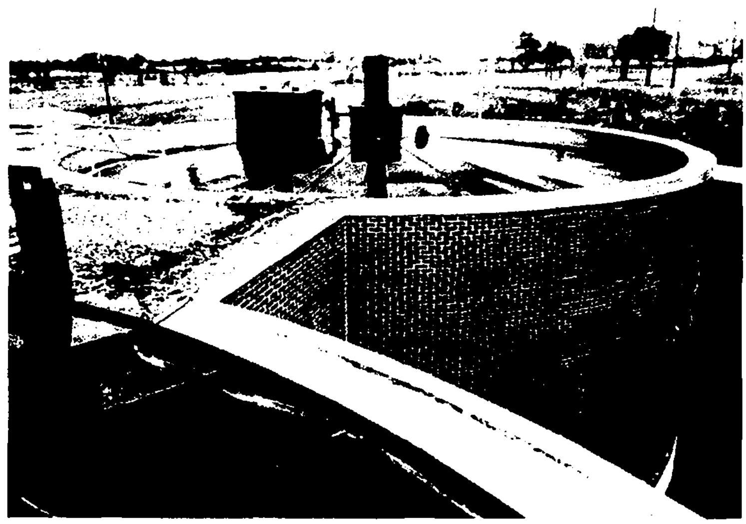


Figure 2–17.—An anaerobic digester floating cover with a gas mixing system.

40 Mgal/d design

600°C − days ÷ 54.4°C = 11 days minimum HRT.

For both designs, a three-day storage capacity also is desired. This dictates that floating covers will be utilized with minimum hydraulic detention time based on when the cover rests on landing corbels and maximum detention time based on when the cover is floating at maximum liquid level.

Figure 2–16 indicates the possible safe range of organic loading for a given HRT.

The practical upper limit on feed solids concentration is 8 to 9 percent.

Within the constraints given, the designer has considerable latitude for selection of digester tank volume (see figure 2–17 for example) and, to a certain point, selection of necessary thickening equipment. For the designs given, the following organic loading has been selected:

4 Mgal/d design—0.15 lbs VS/ft³/day

$$\frac{5{,}800 \text{ lbs VS/day}}{0.15 \text{ lb VS/ft}^3\text{/day}} \times \frac{7.48 \text{ gal}}{\text{ft}^3} \times \frac{1}{17 \text{ day minimum}}$$

$$= 17{,}014 \text{ gallons/day}$$

$$\frac{8{,}000 \text{ lbs solids/day}}{(17{,}014 \text{ gal/day})(8.34)} \times 100$$

$$= 5.64 \text{ percent feed solids required}$$

Minimum tank volume:

17 day × 17,014 gal/day = 289,238 gal (38,668 ft³).

Maximum tank volume:

20 day × 17,014 gal/day = 340,280 gal (45,492 ft³).

Use one digester, 45 ft (13.7 m) diam, 5.7 ft (1.7 m) deep cone, 28.7 ft (8.7 m) side wall depth with 4.3 ft (1.3 m) cover travel.

40 Mgal/d design—0.20 lbs VS/ft³/day

$$\frac{58{,}000 \text{ lbs VS/day}}{0.2 \text{ lb VS/ft}^3\text{/day}} \times \frac{7.48 \text{ gal}}{\text{ft}^3} \times \frac{1}{11 \text{ day minimum}}$$

$$= 197{,}200 \text{ gal/day}$$

$$\frac{80{,}000 \text{ lbs solids/day}}{(197{,}200 \text{ gal/day})(8.34)} \times 100$$

$$= 4.87 \text{ percent feed solids required}$$

Minimum tank volume.

11 day × 197,200 gal/day = 2,169,200 gal (290,000 ft³).

Maximum tank volume.

14 day × 197,200 gal/day = 2,760,800 gal (369,091 ft³).

Use two digesters, each 95 ft diam, 11.9 ft deep cone, 24.5 ft side wall depth with 5.6 ft cover travel.

Table 2–19 gives various calculated results for volatile suspended solids destruction in an anaerobic digester.

## Expected Energy Production

Depending on sludge composition (oil, grease, fiber, protein), gas production can range from 12 to 18 ft³/lb (0.75–1.12 m³/kg) VS destroyed, with the higher values indicating high grease content.

Depending on methane content, each cubic foot of digester gas has an energy value between 550 to 650 Btu (580–685 kJ).

4 Mgal/d design at 55 percent VS destruction.

| *lbs VS destroyed per day* | *Cu ft produced per lb VS destroyed* | *Total cu ft produced per day* | *Btu per cu ft* | *Total Btu produced per day × $10^6$* |
|---|---|---|---|---|
| 3,190..... | 12 | 38,280 | 550 | 21.054 |
| | | | 600 | 22.960 |
| | | | 650 | 24.862 |
| | 15 | 47,850 | 550 | 26.317 |
| | | | 600 | 28.710 |
| | | | 650 | 31.102 |
| | 18 | 57,420 | 550 | 31.581 |
| | | | 600 | 34.452 |
| | | | 650 | 37.323 |

Table 2–19.—Various calculated results for volatile suspended solids destruction in anaerobic digester

| | 4 Mgal/d design | 40 Mgal/d design |
|---|---|---|
| Lbs volatile suspended solids (VSS) destroyed per day | 0.55 (2,040 + 3,760) = 3,190 | 31,900 |
| Percent of TS destroyed | $\frac{3{,}190}{8{,}000} \times 100 = 39.9$ | 39.9 |
| Percent of biodegradable VS destroyed | $\frac{3{,}190}{3{,}760} \times 100 = 84.8$ | 84.8 |
| Percent original inlet feed VSS/TS | $\frac{5{,}800}{8{,}000} \times 100 = 72.5$ | 72.5 |
| Percent final VSS/TS | $\frac{5{,}800 - 3{,}190}{8{,}000} \times 100 = 32.6$ | 32.6 |

40 Mgal/d design at 55 percent VS destruction.

Would be same as 4 Mgal/d except 10 times greater.

Note: 1 hp-hr = 2,545 Btu; electrical energy conversion 32 to 37 percent.

## Sludge Heat Requirements

4 Mgal/d design.—40 hrs/wk.

$$\frac{17{,}014 \text{ gal}}{\text{day}} \times \frac{7 \text{ day}}{\text{wk}} \times \frac{1 \text{ wk}}{40 \text{ hrs}} \times \frac{8.34 \text{ lbs}}{\text{gal}} (95\text{–}40)^\circ\text{F}$$

$$= 1{,}365{,}756 \text{ Btu/hr}$$

40 Mgal/d design.—12 hrs/day—7 day/wk.

$$\frac{197{,}200 \text{ gal}}{\text{day}} \times \frac{1}{2 \text{ units}} \times \frac{8.34 \text{ lbs}}{\text{gal}} \times \frac{(130\text{–}40)^\circ\text{F}}{12 \text{ hrs}}$$

$$= 6{,}167{,}430 \text{ Btu/hr/unit}$$

For both designs, the designer has selected a floating cover with wood composition roof, 12-in. (30.5 cm) thick concrete wall with air space and 4-in. (10.2 cm) brick, and 12-in. (30.5 cm) thick concrete floor exposed to wet earth (table 2-20).

## Summary of Heat Requirements

| *Btu/hr* | *4 Mgal/d* | *40 Mgal/d* |
|---|---|---|
| Expected max. winter output per unit | 1,491,584 | 6,738,006 |
| Expected max. winter input per unit (heat ex. only 80 percent efficient) | 1,864,480 | 8,422,508 |
| Expected total max. winter Btu requirement | $14.9 \times 10^6$ | $219.3 \times 10^6$ |
| Expected total min. summer (air at 75°F, sludge at 50°F) Btu requirement | $10.5 \times 10^6$ | $182.3 \times 10^6$ |
| Expected min. summer input per unit (heat ex. only 80 percent efficient) | 1,482,643 | 7,224,150 |

## Matching Output With Requirement

*4 Mgal/d*

| *Expected total Btu produced per day × 10⁶* | *Average hourly production, Btu/hr* | *Max. req. winter conditions, Btu/hr* | *Min. req. summer conditions, Btu/hr* |
|---|---|---|---|
| 21.054 | 877,250 | | |
| 22.960 | 956,666 | | |
| 24.862 | 1,035,916 | | |
| 26.317 | 1,096,541 | | |
| 28.710 | 1,196,250 | | |
| 31.102 | 1,295,916 | | |
| 31.581 | 1,315,875 | | |
| 34.452 | 1,435,500 | | |
| 37.323 | 1,555,125 | | 1,482,643 |
| | | 1,864,480 | |

Note that the maximum hourly requirement is above the expected hourly production and that even the minimum just makes it, even though total maximum requirements are below minimum total expected gas production. There are three actions which can be taken: (1) Operate digester at lower temperature, (2) increase heat exchanger operating time, and (3) provide some type of gas storage, either a low pressure gas holder (12–24 hr capacity) or high pressure gas holder (several weeks' capacity).

An example of a system for heating anaerobic digesters is shown in figure 2–10.

*40 Mgal/d*

| *Expected total Btu produced per day × 10⁶* | *Average hourly production, Btu/hr* | *Max. req. winter conditions, Btu/hr* | *Min. req. summer conditions, Btu/hr* |
|---|---|---|---|
| 210.54 | 8,772,500 | 8,422,508 | 7,224,150 |
| 229.60 | 9,566,660 | | |
| 248.62 | 10,359,160 | | |
| 263.17 | 10,965,410 | | |
| 287.10 | 11,962,500 | | |
| 311.02 | 12,959,160 | | |
| 315.81 | 13,158,750 | | |
| 344.52 | 14,355,000 | | |
| 373.23 | 15,551,125 | | |

Figure 2–18 shows the general system layout proposed for both designs.

Table 2–20.—Maximum winter—full tank heat radiation loss (Btu/hr)

| | 4 Mgal/d | 40 Mgal/d |
|---|---|---|
| Roof/tank | 44,611 | 251,925 |
| Wall/tank | 71,162 | 252,840 |
| Floor/tank | 10,055 | 65,811 |
| Total | 125,828 | 570,576 |

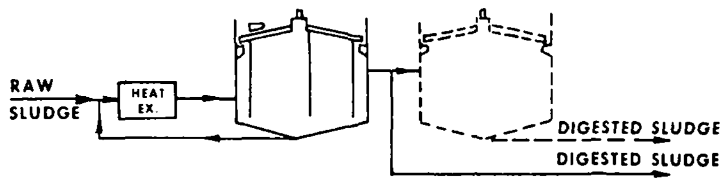



Figure 2–18.—Schematic of conventional high-rate anaerobic digestion system.

## SIZING GAS SAFETY EQUIPMENT

The objective is to remove moisture and convey digester gas from digester to gas utilization, storage or flaring device.

Since hourly production fluctuates greatly each day, it is common to size piping to handle 2.5 times the hourly average.

## 4 Mgal/d Design

Assume all gas is produced in one digester.
Possible to produce 57,420 $ft^3$/day.

$$\frac{57{,}420\ ft^3/day}{24\ hrs/day} \times 2.5 = 5.963\ ft^3/hr$$

$$\frac{5{,}963\ ft^3/hr}{A \times 3{,}600\ sec/hr}$$

A = Cross sectional area inside pipe

| | |
|---|---|
| 2 in. pipe A = 0.022 $ft^2$ | 4 in. pipe A = 0.088 $ft^2$ |
| 3 in. pipe A = 0.049 $ft^2$ | 6 in. pipe A = 0.196 $ft^2$ |

With a 4-in. (10.2 cm) pipe, the maximum velocity is almost 19 ft/sec (5.8 m/s), well in excess of the 11 to 12 ft/sec (3.4–3.7 m/s) recommended for successful condensate removal. Rather than increase the line size to 6 in (15.2 cm), it is recommended that oversize accumulators be used.

Gas safety piping specifications are as follows (see figure 2–9):

1. All gas lines must be tight, sloped (1/4 in./ft)(2.1 cm/m) toward condensate traps and accumulators, have ample capacity and be protected against freezing.
2. Lines leading to gas burners or gas engines must be protected against flashbacks by flame traps. Trap should be located near point of combustion with a maximum allowable distance of 30 ft (9.1 m) from point of gas combustion.
3. Bypasses are needed to permit flexibility of operation, but flame traps are never bypassed.
4. Total pressure loss through the appurtenances and gas lines from the digester to use should be only 2.0 in. (5.1 cm) W.C. at maximum gas flow rate.

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# Aerobic Digestion and Design of Municipal Wastewater Sludges

Aerobic digestion of municipal wastewater sludges is based on the principle that with inadequate external food sources, biological cells will consume their own cellular material.

The claimed advantages of aerobic digestion are:[1]

- Volatile solids reduction approximately equal to that obtained anaerobically.
- Low $BOD_5$ concentrations in the supernatant liquor.
- Production of an odorless, humuslike, biologically stable end product that can be disposed of easily.
- Production of a sludge with excellent dewatering characteristics.
- Recovery of more of the basic fertilizer value in the sludge.
- Few operational problems.
- Low capital cost.

Table 3–1 indicates the types of sludges that have been studied on a full-scale basis. Operating results from operating installations indicate that the third and fourth items above are not correct. Some aerobically digested sludges, even at very long detention times are not biologically stable and aerobically digested sludge does not dewater very readily with mechanical equipment.

Today most aerobic digesters are designed using rules of thumb developed from past experience (table 3–2) and as the literature has noted[19–22] do not always perform as intended. This chapter presents the most up-to-date design criteria available. Whenever possible full-scale operating data are presented.

Table 3–1.—Type and reference of full-scale studies on aerobic digestion of municipal wastewater sludge

| | Reference on mesophilic | Reference on thermophilic |
|---|---|---|
| Primary sludge only | 2,3,4,5,18 | 6 |
| Waste activated only | 7,8 | 9,10 |
| Mixed primary and waste activated sludge | 7,8,11 | 10,12,13 |
| Waste activated sludge from contact stabilization | 14,15 | |
| Primary and lime | 16 | |
| Trickling filter only | 2 | |
| Mixed primary and trickling filter | 2 | |
| Sludges containing iron or alum | 17,18 | |

Table 3–2.—Typical present day aerobic digestion design criteria[23]

| 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, lbs volatile solids/$ft^3$/day | 0.1–0.20 |
| Oxygen requirements, lb/lb cell destroyed | [b]2.0 |
| Energy requirements for mixing | |
| Mechanical aerators, hp/1,000 $ft^3$ | 0.5–1.0 |
| Air mixing, scfm/1,000 $ft^3$ | 20–30 |
| Dissolved oxygen level in liquid, mg/l | 1–2 |

[a]Detention times should be increased for temperatures below 20°C. If sludge cannot be withdrawn during certain periods (e.g., weekends, rainy weather) additional storage capacity should be provided.

[b]Ammonia produced during carbonaceous oxidation oxidized to nitrate.

## CRYOPHILIC—MESOPHILIC—THERMOPHILIC DIGESTION

For purposes of classification the following three temperature zones of bacterial action will be used throughout this chapter:

- Cryophilic zone—liquid temperature below 10°C (50°F)
- Mesophilic zone—liquid temperature between 10°C to 42°C (50°F to 108°F)
- Thermophilic zone—liquid temperature above 42°C (108°F)

The effect of temperature on the effectiveness of aerobic digestion is still an area of considerable controversy,[24] especially in the areas of solids reduction, dewaterability and settleability. The data shown in subsequent sections should help clarify some of the controversy.

At the present time considerable research is being undertaken in the design and operation of thermophilic aerobic systems,[13,24–31] especially auto-thermophilic aerobic systems.[13,27,29,31] Claimed advantages of the thermophilic aerobic system are:[13,30,31]

- Higher rates of organic stabilization that allow smaller volume requirements.
- Higher maintenance energy requirements and higher microbial decay rates that give smaller amounts of sludge for disposal.
- Digestion in this temperature range should make liquid essentially pathogen free.
- All weed seeds should be destroyed.
- Total oxygen demand should be 30 to 40 percent less than mesophilic since few, if any, nitrifying bacteria exist in this temperature range.
- Improved solids-liquid separation due to decreasing liquid viscosity.
- Possible improved oxygen transfer rates because of the significantly higher coefficient of diffusivity of oxygen.

## VOLATILE SOLIDS REDUCTION

One of the main objectives of aerobic digestion is to reduce the amount of solids that need to be disposed. This reduction is normally assumed to take place only with the volatile content of the sludge, though some studies[24,32] have shown that there can be destruction of the nonorganics as well. In this discussion solids reduction will pertain only to the volatile content.

The change in volatile content is normally represented by a first order biochemical reaction,

$$dx/dt = -K_d X \qquad (1)$$

where

$dx/dt$ = rate of change of volatile suspended solids per unit of time
$K_d$ = reaction rate constant – day$^{-1}$
$X$ = concentration of volatile suspended solids at time t in aerobic digester

The time t in equation (1) is actually the sludge age in the aerobic digester and, depending on how the aerobic digester is being operated (continuous flow without recycle or with recycle, batch with supernatant decant), can be considerably greater than the theoretical hydraulic residence time (HRT).

A distinction must be made between biodegradable volatile suspended solids and nonbiodegradable volatile suspended solids. Research in this area is quite limited but the following generalities can be used.

- Approximately 20 to 30 percent of the influent suspended solids of a typical domestic wastewater is inert.[33] Of the remaining suspended solids that are volatile, approximately 40 percent are inert organics consisting chiefly of lignins, tannins, and other large complex molecules.

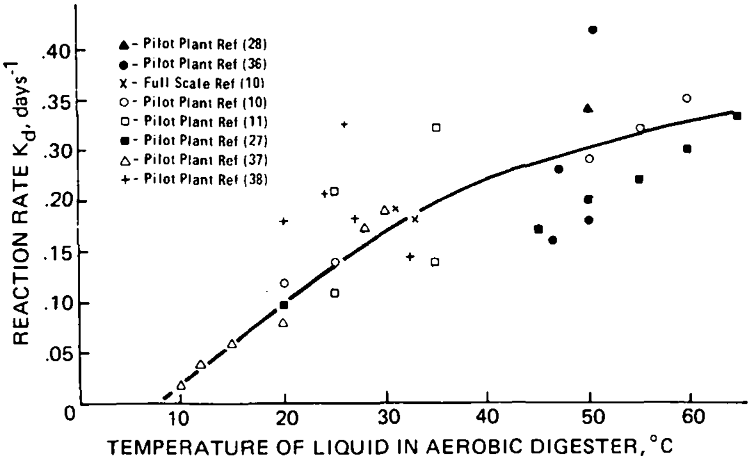


Figure 3–1.—Reaction rate $K_d$ versus liquid temperature in digester.

- For waste-activated sludges generated from primary treatment systems, approximately 20 to 35 percent of the volatile solids produced are nonbiodegradable.[34,35]
- For waste-activated sludges generated from the contact-stabilization process (no primaries—all influent flow into aeration tank), 25 to 35 percent of the volatile suspended solids are nonbiodegradable.[15]

The reaction rate term $K_d$ is a function of sludge type, temperature and solids concentration. It is only a pseudo constant since the term actually is the average result of the many variables affecting it at any one time. Figure 3–1 shows a plot of various reported $K_d$ values as a function of the liquid temperature in the aerobic digester. The data shown are for several types of waste sludge, which probably is a partial reason for the scatter. At

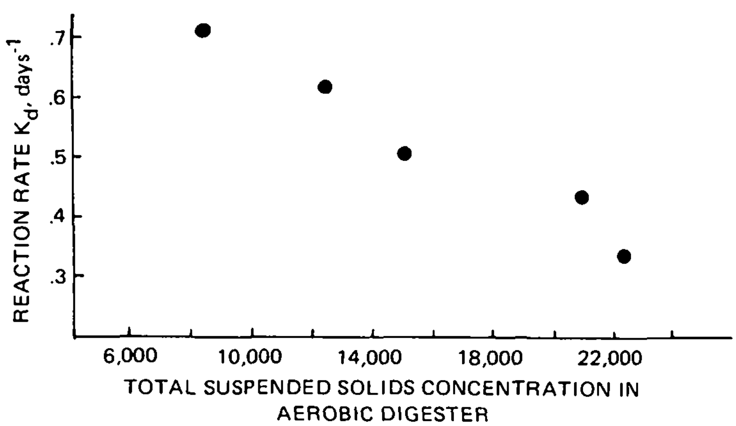


Figure 3–2.—Effect of aerobic digester solids concentration on reaction rate $K_d$ done at 20° C waste-activated sludge.

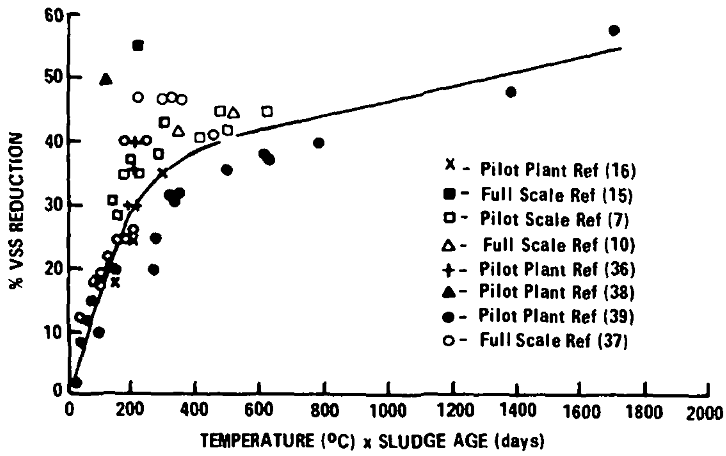


Figure 3–3.—Volatile suspended solids reduction as a function of digester liquid temperature and digester sludge age.

this time there is not enough data to allow segregation of $K_d$ by sludge type; therefore, the line drawn through the data points represents an overall average $K_d$ value.

Figure 3–2 indicates the results from one study[15] on the effects of aerobic digester solids concentration on the reaction rate, $K_d$. Figure 3–3 shows the effect of temperature and sludge age on total volatile suspended solids reduction.

## OXYGEN REQUIREMENTS

Activated sludge biomass is most often represented by the empirical equation $C_5H_7NO_2$. Under prolonged periods of aeration, typical of the aerobic digestion process, the biochemical equation for oxidation is represented by equation (2).

$$C_5H_7NO_2 + 7O_2 \rightarrow 5CO_2 + 3H_2O + H^+ + NO_3^- \qquad (2)$$

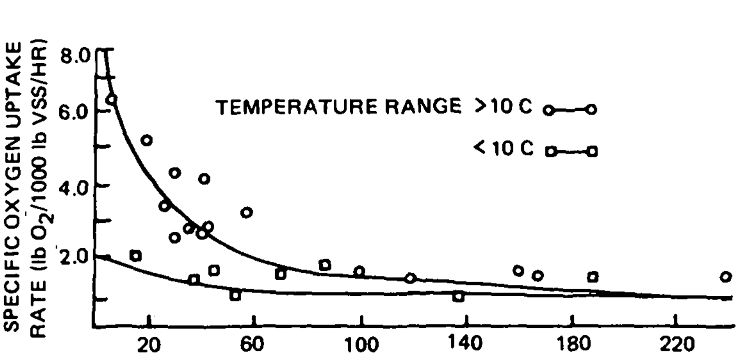


Figure 3–4.—Effects of sludge age and liquid termperature on oxygen uptake rates in aerobic digesters.[19]

Theoretically, this reaction states that 1.98 pounds of oxygen are required per pound of cell mass oxidized. In pilot[36] and full-scale[10,15] studies where this value has been evaluated, the range was from 1.74 to 2.07 pounds of oxygen required per pound of volatile solids destroyed. For mesophilic systems, a design value of 2.0 is recommended. For thermophilic systems where nitrification would not exist,[13,30,31] a value of 1.4 is recommended.

The actual specific oxygen utilization rate, pounds of oxygen per 1,000 pounds volatile solids per hour, is a function of total sludge age and liquid temperature.[19,24,38] In one study, Ahlberg and Boyko[19] visited several operating installations and developed the relationship shown in figure 3–4. Field studies[19] have indicated that a minimum value of 1.0 mg of oxygen should be maintained in the digester at all times.

## MIXING

Mixing in an aerobic digester, treating municipal wastewater sludge of domestic origin, is considered to have the following benefits. (Note: It is assumed that a favorable environment exists to allow development of an aerobic digestion system.)

- It continues to bring deoxygenated liquid to the aeration device.
- It keeps the food supply uniformly dispersed and in constant contact with the growing cells to promote maximum utilization of the system.
- It keeps the concentration of biological end products at their lowest value by dispersing them uniformly throughout the digester.
- It provides environmental uniformity (oxygen, temperature, nutrients, etc.) throughout the digester to allow the best possible cell development.
- It allows fairly fast dispersion of any toxic material entering the system, thus, possibly minimizing its effect on the aerobic process.

There is general agreement that mixing is an important criterion in the aerobic digestion process. The problem arises when one tries to evaluate, define or specify a mixing system.

In recent years it has become popular to use the term "complete mix" when discussing biological process reactors. The term "complete mix" means that the time for dispersion of the feed stream is short in relation to the total hydraulic residence time in the reactor. It is also defined as sufficient mixing so that concentration gradients of chemical and biological ingredients are uniform for the particular reaction rates that exist in the basin.

Mixing within the aerobic digestion tank occurs on two levels: macromixing and micromixing.[43] Macromixing deals with the bulk mass flow within the digester, while micromixing deals with the degree of intermingling of the system molecules. In biological theory, "complete mix" assumes micromixing.[44]

The actual mixing can be performed by a gas system, mechanical system or a combination of the two.

No matter what type device is utilized, the intent is to achieve mixing through a pumping action. Because of this relationship, engineers have come to use the term horsepower/unit volume as some type of parameter to define mixing in an aerobic digester. Unfortunately, this term by itself has no meaning. For mechanical type mixers the wide variation in impeller diameters and speeds can result in similar horsepower but widely different pumping capacities. For gas mixing systems gas flow, depth, and bubble size can also result in similar horsepower but widely different pumping capacities. In addition, tank geometry and solids concentrations can significantly affect power requirements.

Probably the best way to define mixing is from the standpoint of zone of influence of an energy source (figure 3–5). Essentially the zone of influence states that energy is dissipated as one moves horizontally away from the energy source. This loss is due to friction between the fluid molecules which is a function of liquid density, temperature, and solids concentration. Within a certain area of the point source there is sufficient energy to achieve micromixing. There is also a larger area where bulk flow (macromixing) still takes place even though there is insufficient energy for micromixing.

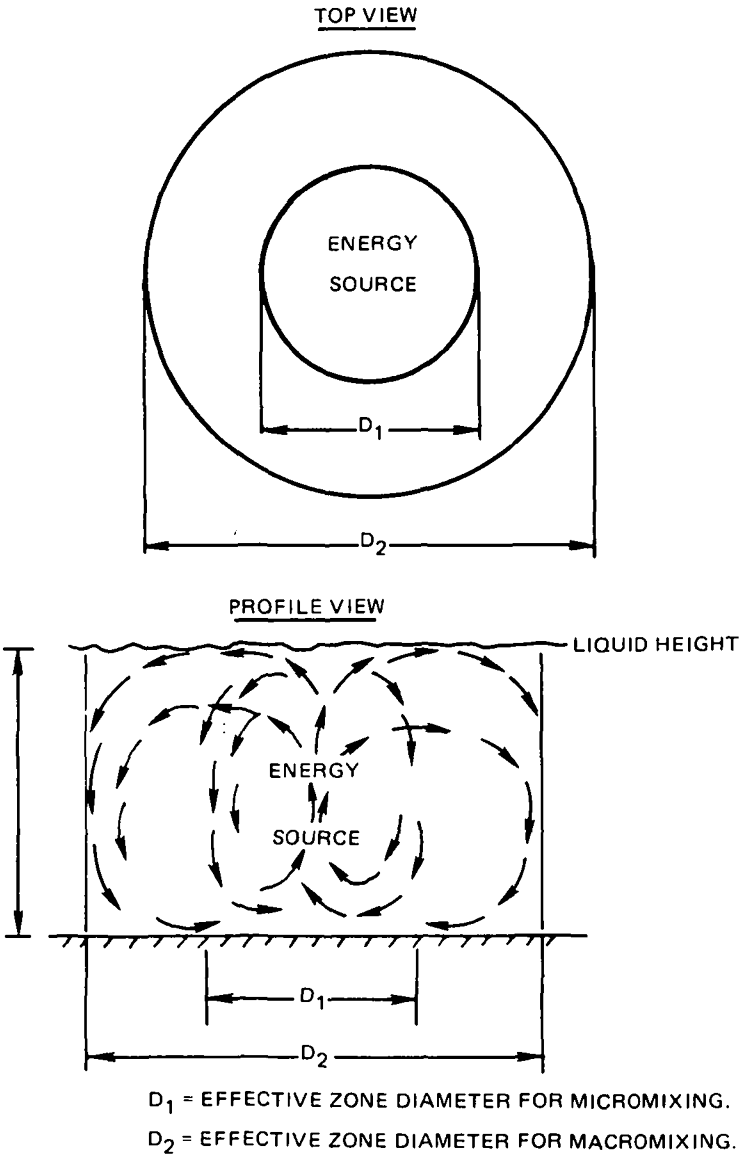


Figure 3–5.—Shear-stress relationship for a thixotropic pseudo plastic material.

Studies[45,46] done with point energy sources, in clean water and with no side boundaries (only surface and floor boundaries) have indicated that the width of the micromixing zone is no more than twice the liquid depth, with the liquid depth being a function of the type of mixing device utilized and not necessarily the tank liquid depth.

The effect of tank geometry[47] on mixing (as measured by oxygen transfer rates in clean water) for various aeration devices (high and low speed mechanical aerators, submerged turbines, oxidation ditch aerator and diffused aeration) in tanks from several thousand to 1 million gallons (~10 to 3,800 $m^3$), was shown to fall into three general categories (figure 3–6).

Category 1 is represented by basin geometry A in figure 3–6. This is the idealized case in which geometry has no effect on the liquid flow pattern. Each increment of power into this specific volume has a corresponding increase in the oxygen supplied.

Category 2 is represented by basin geometry B in figure 3–6 and has been termed the "flywheel effect."

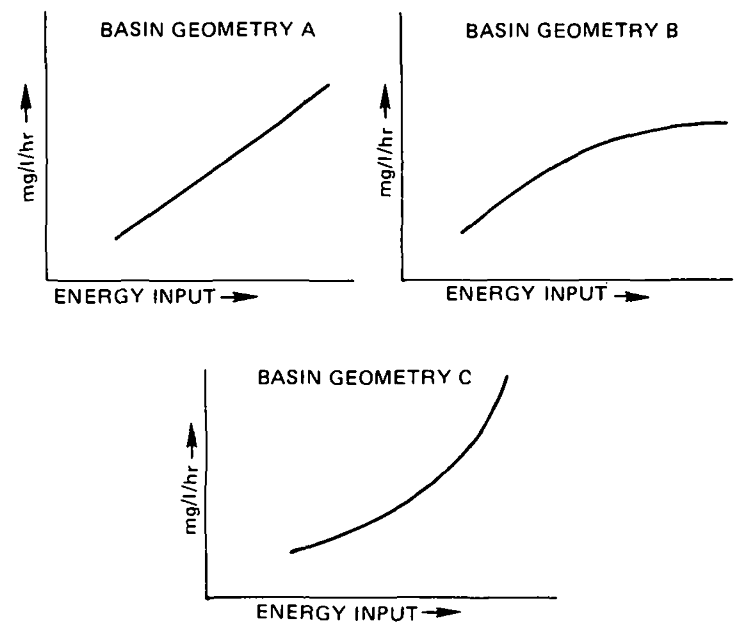


Figure 3–6.—Schematic of zone of mixing influence for energy source in fluid with only fixed upper and lower boundaries.

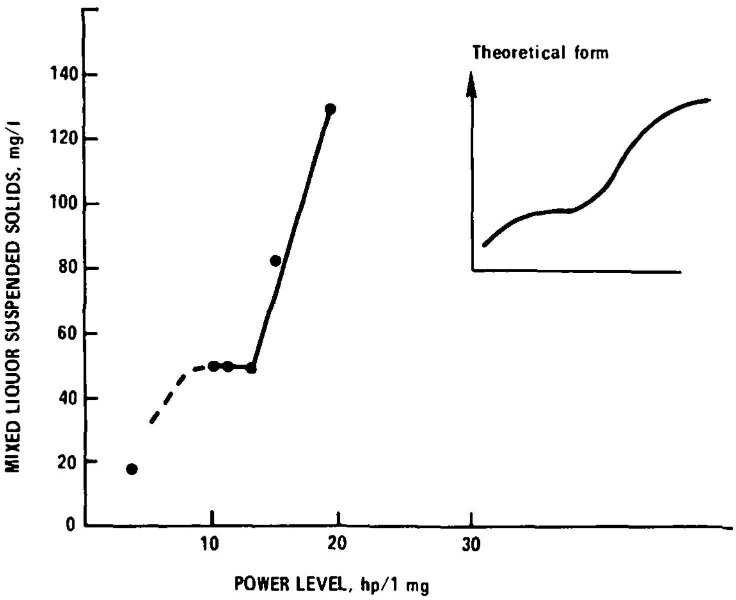


Figure 3–7.—Effects of tank geometry on mixing in clean water as measured by oxygen transfer rates.

Here tank constraints, represented, for example, by a channel aeration tank, cause a rapid increase in oxygen supply for small inputs of energy. As the energy per unit volume increases, the geometry of tanks causes a leveling off in transfer.

Category 3 is represented by basin geometry C in figure 3–6 and has been termed the "choke flow effect." Here tank geometry interferes with the mixing pattern until a certain energy level is reached. At this point there is sufficient energy to override the constraint and allow for complete mixing in the tank contents.

No published studies on field evaluation of the effect of suspended solids on mixing in aerobic digesters are available. There have been several such studies[48,50] conducted in lagoons with suspended solids in the range of 100 to 400 mg/l and figure 3–7 shows the results. In general, increased solids concentrations required increased power levels, though the tank geometry[50] and interaction effects of other aerators[49] also influenced mixing patterns.

## CHARACTERISTICS OF AEROBIC DIGESTERS

The existing trend in wastewater treatment is to remove more and more material from the main liquid processing stream. This is frequently done through the use of secondary biological treatment schemes, chemical treatment and filtration. The sludge produced can vary widely and change rapidly even on an hour-to-hour basis.

Table 3–3 gives specific gravity and particle size distribution on two common type sludges: plain primary and plain waste activated.

Table 3–3.—General characteristics of raw primary and waste activated sludge[40]

| | Primary sludge | Waste activated sludge |
|---|---|---|
| Specific gravity | 1.33–1.4 | 1.01–1.05 |
| Particle size | 20% <1 μm | 40% 1–50 μm |
| | 35% 1–100 μm | 60% 50–180 μm |
| | 45% <100 μm | |
| Physical appearance | Fibrous | Slimy, gelatinous |

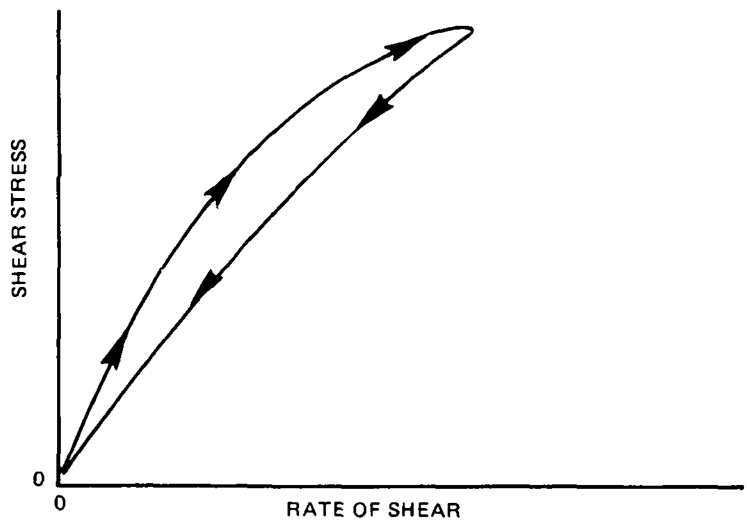


Figure 3–8.—Power level versus suspended solids.[50]

There is little data on the rheology of municipal wastewater sludge,[40] and none could be found on strictly aerobically digested sludge. One of the main problems with collecting data is that such studies are extremely difficult to perform correctly.[41]

Even though the majority of raw wastewater sludges behave as a thixotropic (time dependent), pseudo plastic material (figure 3–8), it may not be correct to assume that the sludge within the aerobic digester has the same general properties. The liquid will have a variable solids concentration and there is a general reduction in particle size and shape,[38,43] both of which affect fluid viscosity.

Another characteristic of present-day designs is that the tanks tend to have large surface area to liquid depth ratios.

## SUPERNATANT

It is common practice in most aerobic digestion facilities not to prethicken the sludge but to concentrate it

Table 3–4.—Characteristics of mesophilic aerobic digester supernatant

| | Reference 9[a] | Reference 19[b] | Reference 52[c] |
|---|---|---|---|
| Turbidity | 120 | — | — |
| $NO_3$-N | 40 | | 30 |
| TKN | 115 | 2.9–1,350 | |
| COD | 700 | 24–25,500 | |
| $PO_4$-P | 70 | 2.1–930 | 35 |
| Soluble | | .4–120 | |
| $BOD_5$ | 50 | 5–6,350 | 2–5 |
| Filtered $BOD_5$ | | 3–280 | |
| Suspended solids | 300 | 9–41,800 | 6.8 |
| Alk | | | 150 |
| $SO_4$ | | | 70 |
| Silica | | | 26 |
| pH | 6.8 | 5.7–8.0 | 6.8 |

[a]Average of 7 months of data.
[b]Range taken from 7 operating facilities.
[c]Average values.

after digestion. This is done by sending the flow to a clarifier-thickener or by turning off the aeration device within the digester for 12 to 18 hours. When this is done, a digester supernatant is taken off which is normally returned to the head end of the treatment plant. Table 3–4 gives supernatant characteristics from several full-scale facilities operating in the mesophilic temperature range.

## pH Reduction

Figure 3–9 shows the effect of sludge age on digester pH for mesophilic operation.

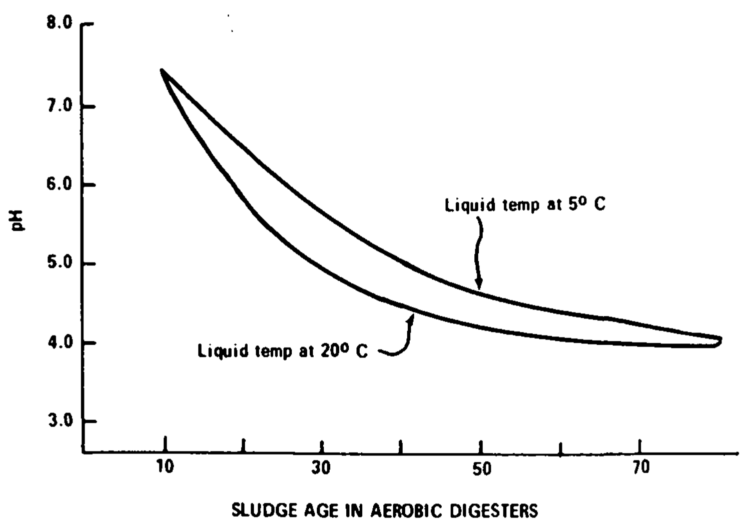


Figure 3–9.—Effects of sludge age on pH for mesophilic aerobic digestion.

The drop in pH is caused by an increased concentration of nitrate ions and a corresponding loss of alkalinity due to the conversion of $NH_3$-N to $NO_3$-N commonly called nitrification. Though at one time, the low pH was considered inhibitory to the process, it has been shown that over time the system will acclimatize and perform just as well at the lower pH values.[7,38,51]

It should be noted that if nitrification does not take place, there will be very little, if any, pH drop. This could happen at low liquid temperatures and short sludge ages or in thermophilic operation.[31] Nitrifying bacteria are sensitive to heat and do not exist in temperatures over 45° C.[52]

## BACTERICIDAL EFFECTS

Pathogenic organisms in wastewaters consist of bacteria, virus, protozoa and parasitic worms; a good current review on the subject can be found in Kenner et al.[56] Many of these organisms, especially enteric viruses,[54] have a strong tendency to bind themselves to sludge solids.

Table 3–5.—Human enteric pathogens occurring in wastewater and the diseases associated with the pathogen[56]

| Pathogens | Diseases |
|---|---|
| Vibrio cholera | Cholera |
| Salmonella typhi | Typhoid and other enteric fevers |
| Shigella species | Bacterial dysentery |
| Coliform species | Diarrhea |
| Pseudomonas species | Local infection |
| Infectious hepatitus virus | Hepatitis |
| Poliovirus | Poliomyelitis |
| Entamoeba histolytica | Amoebic dysentery |
| Pinworms (eggs) | Asceariasis |
| Tapeworms | Tapeworm infestation |

Table 3–6.—Pathogenic organisms in sludge[57]

| Type | Salmonella (No./100 ml) | Pseudomonas aeruginosa (No./100 ml) | Fecal coliform (No. × $10^6$/100 ml) |
|---|---|---|---|
| Raw primary | 460 | $46 \times 10^3$ | 11.4 |
| | 62 | 195 | |
| Trickling filter | 93 | $110 \times 10^3$ | 11.5 |
| Raw waste activated sludge | 74 | $1.1 \times 10^3$ | 2.8 |
| | 2,300 | $24 \times 10^3$ | 2.0 |
| | 6 | $5.5 \times 10^3$ | 26.5 |
| Thickened raw waste activated sludge | 9,300 | $2 \times 10^3$ | 20 |

Table 3–7.—Thermophilic aerobic digestion time required for reduction of pathogenic organisms below minimum detectable level[61]

| Type | Temperature °C | Time required for lowest detectable limit of *salmonella* hours | Time required for lowest detectable limit of *pseudomonas aeruginosa* hours |
|---|---|---|---|
| Mixture of primary and waste activated ...... | 45 | 24 | 24 |
| | 50 | 5 | 2 |
| | 55 | 1 | 2 |
| | 60 | 0.5 | 0.5 |

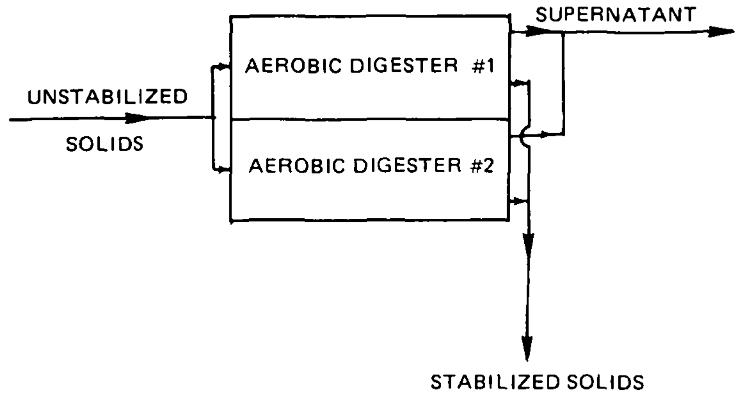


Figure 3–10.—Tank configuration for a batch operated aerobic digester.

Table 3–5 gives a listing of human enteric pathogens occurring in wastewater sludges along with the diseases associated with them. Table 3–6 gives some data on bacterial concentrations of various types of raw sludges.

Researchers have studied pathogenic organism reduction in both mesophilic[56,58,59] and thermophilic digestion.[60] Under mesophilic operation, the bactericidal effects appear to be related to natural die-off with time. For thermophilic operation, the time required for reduction of pathogenic organisms below minimal detection level is a function of basin liquid temperature (table 3–7).

## DEWATERING

One of the supposed benefits of aerobic digestion is the production of a sludge with excellent dewatering characteristics.[1] Much of the published literature on full-scale operations has indicated this is not true,[3,4,17,26,61] though there are some published reports of excellent operating systems.[15]

Although most recent investigators agree that there is a deterioration in dewaterability with increasing sludge age,[2,16,17,27,62] there is still debate as to the cause; lack of sufficient oxygen[26,27] reduction in particle size[16,17] or concentration of biological anionic polymers.[63]

At this time it can only be recommended that conservative design be used for designing mechanical sludge dewatering facilities unless pilot plant data indicate otherwise.

## TANK LAYOUTS AND OPERATION

Originally aerobic digesters were operated as strictly a batch operation and this concept is still used at many facilities (figure 3–10).

Solids are pumped directly from the clarifiers into the aerobic digester. Eventually, the digester fills up, and the time required depends not only on the waste sludge flow but on the amount of precipitation or evaporation. When the tank is full, the aeration device is turned off for several hours to allow solids-liquid separation, then a decant operation takes place. After decanting, thickened stabilized solids of about 2 to 4 percent in concentration, can be removed and more waste sludge can be added.

Many engineers tried to make the process more continuous by installing stilling wells in part of the digester. This has proved not to be effective[20,64,65] and should not be incorporated into the design.

The next step was then to provide the aerobic digester with its own clarifier-thickener (figure 3–11).

Solids are still pumped directly from the clarifiers into the aerobic digester. In this case the aerobic digester operates at a fixed level with the overflow going to a solids-liquid separator. Thickened solids are normally recycled back to the digestion tank but when required can also be removed from the system.

Though initially more costly than a batch operated system, much of the manual work involved with aerobic digestion is eliminated.

A third type of system would involve prethickening before aerobic digestion. This is employed in the currently being researched auto thermophilic aerobic digestion system (figure 3–12).

In this system, sludge from the clarifiers would go to some type of thickening device to produce a concentration greater than 4 percent solids then into the digester. When operating in this mode, one should not expect any

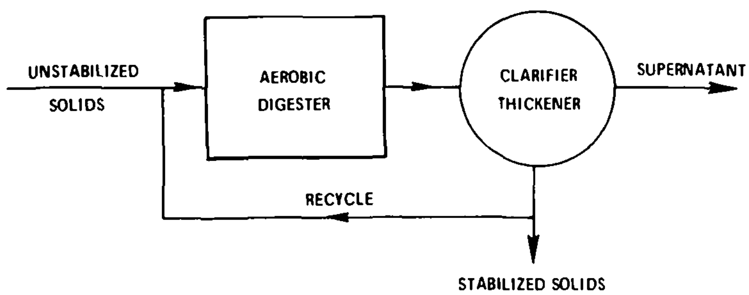


Figure 3–11.—Tank configuration for a continuous operated aerobic digester.

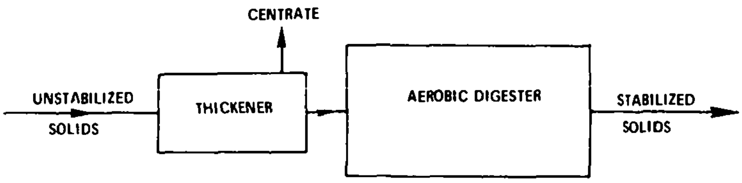


Figure 3–12.—Tank configuration for an auto thermophilic aerobic digestion system.

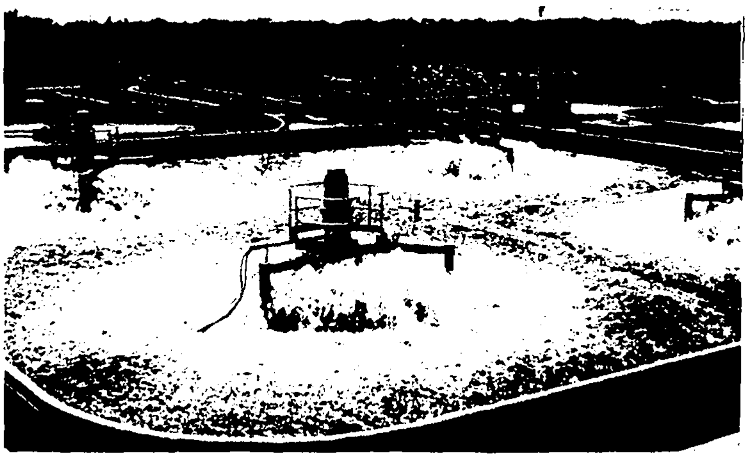

Figure 3–13.—A floating low speed aerator in an aerobic digester.

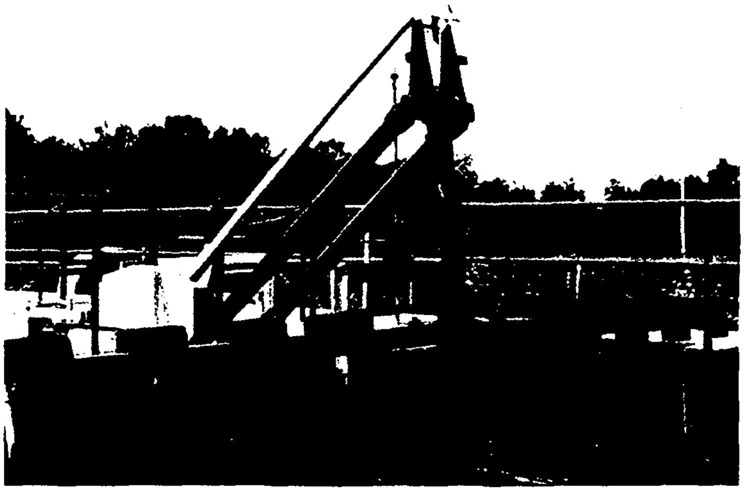

Figure 3–14.—A diffused air system in an aerobic digestion tank.

further gravity solids-liquid separation to take place after digestion (see figures 3–13 and 3–14).

## SUMMARY

The basic design criteria for aerobic sludge digestion systems presented in the previous sections are summa-

Table 3–8.—Criteria for design of aerobic digestion systems

| | Days | Liquid temperature |
|---|---|---|
| Sludge age required to achieve | | |
| 40 percent volatile solids reduction | 108 | 4.4°C (40°F) |
| | 31 | 15.5°C (60°F) |
| | 18 | 26.7°C (80°F) |
| 55 percent volatile solids reduction | 386 | 4.4°C (40°F) |
| | 109 | 15.5°C (60°F) |
| | 64 | 26.7°C (80°F) |
| Oxygen requirements | Liquid temperature 45°C or less; 2.0 lbs. oxygen/lb volatile solids destroyed<br>Liquid temperature greater than 45°C; 1.4 lbs. oxygen/lb volatile solids destroyed | |
| Oxygen residual | 1.0 mg $O_2$/l at worst conditions | |
| Expected maximum solids concentration achievable with decanting | 2.5 to 3.5 percent solids (degritted sludge) | |
| Mixing horsepower | Function of tank geometry and type of aeration equipment utilized | |

rized in table 3–8. Obviously, operational criteria will vary with the quantity and biodegradability of material to be stabilized, as well as temperature and other critical parameters.

## DESIGN PROBLEM

Two designs, a 4 Mgal/d (.18 $m^3$/s) and 40 Mgal/d (1.75 $m^3$/s), are evaluated. Influent is typical domestic wastewater of 200 mg/liter biochemical oxygen demand ($BOD_5$) and 200 mg/liter suspended solids (SS) with no heavy industrial contributors. Liquid treatment consists of grit removal, primary treatment, secondary treatment (activated sludge) and chlorination. No chemicals are added to liquid treatment portion.

### Sludge Type and Amount

Every million gallons (3,785 $m^3$) of raw plant influent will generate approximately 1,000 lbs. (453.6 kg) of dry primary sludge and 1,000 lbs. (453.6 kg) of waste-activated sludge solids.[67] Table 3–9 shows how this can be further broken down.

Based on table 3–9 the sludge generated for the two design examples would be

| | *4 Mgal/d design (lbs)* | *40 Mgal/d design (lbs)* |
|---|---|---|
| Inert nonvolatile | 4×550 = 2,200 | 40×550 = 22,000 |
| Inert volatile | 4×510 = 2,040 | 40×510 = 20,400 |
| Biodegradable volatile | 4×940 = 3,760 | 40×940 = 37,600 |
| Total | 4×940 = 8,000 | 40×940 = 80,000 |

Table 3–9.—Breakdown of inert and volatile suspended solids per mg of plant influent (lbs)

| | Inert nonvolatile | Inert volatile | Biodegradable volatile |
|---|---|---|---|
| Primary sludge | 250 | 300 | 450 |
| Waste activated sludge | 300 | 210 | 490 |
| Totals | 550 | 510 | 940 |

## Temperature Effect

Temperature in the aerobic digestion process:

- Affects oxygen transfer capabilities.
- Affects volatile destruction capabilities.

Temperature in aerobic digester is a function of:

- Feed solids concentration.
- Geographical location of treatment facility.
- Tank location and material of construction.
- Type of aeration device utilized.

For this design example the following assumptions will be made:

- Thermophilic or auto-thermophilic aerobic digestion will not be considered. This implies average inlet feed solids to digester under 3.5 percent solids concentration.
- Lowest liquid temperature expected during winter is 10°C (50°F). During the summer 25.5°C (78°F) is expected.

## Expected Type of Volatile Solids Destruction

Figure 3–3 showed a plot of volatile suspended solids destruction as a function of liquid temperature and sludge age. A minimum of 40 percent VSS reduction has been chosen for the design example which would require a temperature-sludge age combination of 475 days. At the minimum liquid temperature of 10°C., this would imply a sludge age of 47.5 days. If the system is designed to maintain a 47.5-day sludge age, then during the summer this combination would be $47.5 \times 25.5 = 1211$ °C-days. This would give a 49 percent reduction. Table 3–10 gives various ratios which could be developed.

## Expected Suspended Solids Concentration in Aerobic Digester Underflow

This is a function of overall detention time, local evaporation rate and type of aerobic digestion system employed (batch or continuous).

Aerobically digested sludge, typically degritted with no chemical addition, can be gravity thickened to 2.5 to 3.5 percent. For this design a maximum of 3.0 percent is assumed.

If there is no prior thickening of the raw sludges so that the average inlet feed solids concentration is under 3.0 percent, then gravity thickening is possible. For this example, the inlet feed solids concentration for the combined sludge is assumed to be 1.5 percent solids (based on 4 percent sludge from the primary clarifier and 1 percent sludge from the secondary clarifier).

## Oxygen Requirements

Since it is assumed that these design examples would not be designed for thermophilic aerobic digestion, nitrification oxygen demand must be met. From previous discussions and for design purposes, 2.0 lbs of oxygen will be considered as the amount required to oxidize a pound of cell mass (table 3–11).

## Minimum Tank Volume Necessary To Achieve Desired Results

It was previously noted that a minimum volatile suspended solids reduction of 40 percent was required at the 10°C liquid level. Based on figure 3–3 this would imply a minimum sludge age ot 47.5 days.

Sludge age in aerobic digester can be approximated as follows:

$$\text{Sludge age} = \frac{\text{total lbs SS in aerobic digester}}{\text{total lbs SS lost per day from aerobic digester}}$$

$$= \frac{\text{total lbs SS in aerobic digester}}{(\text{total lbs SS lost per day in supernatant}) + (\text{total lbs SS wasted per day from system})}$$

$$= \frac{(\text{SS conc. in digester})(8.34)(\text{digester tank volume})}{[(\text{SS conc. in supernatant})(1 - f) + (\text{SS conc. in underflow})(f)]\,(8.34)(\text{influent flow})}$$

where:

$$f = \frac{(\text{influent SS conc.})(\text{percent solids not destroyed})}{\text{thickened SS conc.}}$$

SS conc. in supernatant—if good solids liquid separation takes place can expect about 300 mg/l SS in supernatant.

Table 3–10.—Various calculated results for volatile suspended solids destruction in aerobic digester

| | 4 Mgal/d design | 40 Mgal/d design |
|---|---|---|
| Lbs volatile suspended solids (VSS) destroyed per day | | |
| Winter | 0.4 (2,040 + 3,760) = 2,320 | 23,200 |
| Summer | 0.49 (2,040 + 3,760) = 2,842 | 28,420 |
| Percent of total solids destroyed | | |
| Winter | $\frac{2,320}{8,000} \times 100 = 29\%$ | 29% |
| Summer | $\frac{2,842}{8,000} \times 100 = 35.5\%$ | 35.5% |
| Percent of biodegradable VS destroyed | | |
| Winter | $\frac{2,320}{3,760} \times 100 = 61.2\%$ | 61.2% |
| Summer | $\frac{2,842}{3,760} \times 100 = 75.5\%$ | 75.5% |
| *Original* inlet feed VSS/TS | $\frac{5,800}{8,000} \times 100 = 72.5\%$ | 72.5% |
| Final VSS/TS | | |
| Winter | $\frac{5,800 - 2,320}{8,000} \times 100 = 43.5\%$ | 43.5% |
| Summer | $\frac{5,800 - 2,842}{8,000} \times 100 = 36.9\%$ | 36.9% |

Table 3–11.—Average pounds of oxygen required per day for aerobic digestion system

| | 4 Mgal/d design | 40 Mgal/d design |
|---|---|---|
| Winter | 2.0 × 2,320 = 4,640 | 46,400 |
| Summer | 2.0 × 2,842 = 5,684 | 56,840 |

SS conc. in digester—can range from a minimum equal to the influent SS concentration to a maximum equal to the thickened concentration (assume no evaporation). Assume that on the average SS conc. equal to 70 percent of the thickened concentration.

Digester tank volume—million gallons.

For 4 Mgal/d design

Sludge age = 47.5 days

SS conc. in digester = (0.7)(30,000 mg/l)

SS conc. in supernatant = 300 mg/l

SS conc. in underflow = 30,000 mg/l

$$f = \frac{(1.5\%)(.71)}{3.0\%} = 0.35$$

$$\text{Influent flow} = \frac{8,000}{(0.15)(8.34)}$$

$$= 63,950 \text{ GPD}$$

$$= 0.06395 \text{ Mgal/d}$$

$$47.5 - \frac{(0.7)(30,000)(\text{tank vol})}{(300)(1 - .35) + (30,000)(.35)(0.06395)}$$

$$= \frac{(21,000 \text{ tank vol.})}{697}$$

$$\text{Digester tank volume} = \frac{(697)(47.5)}{21,000} = 1.576 \text{ mg}$$

Tank geometry function of site location and type of aeration equipment to be utilized.

For 40 Mgal/d design

Everything the same except for influent flow which = 0.6395.

Tank volume = 15.76 mg.

In addition to the tank volume calculated, additional volume may be required depending on local weather conditions and type of downstream sludge-handling facilities.

## Tank Layout

For the mesophilic aerobic digestion system being considered, there are two types of systems to choose from: the batch operated system (figure 3–10) or the continuous flow through system (figure 3–11).

The original aerobic digestion systems were batch operated; this is still the most prevalent design (figure 3–10).

Solids are pumped directly from the clarifiers into the aerobic digester. The time required for the tank to fill up depends not only on the waste sludge flow but the amount of precipitation or evaporation. When the tank is full, the aeration device is turned off for several hours to allow solids-liquid separation, then a decant operation takes place. After decanting, thickened stabilized solids, about 3 percent, can then be removed or more waste sludge would be added.

In the past, many engineers have tried to make this design more continuous by installing stilling wells in part of the tank. This has proved not to be effective[20,64,65] and should not be incorporated into the design.

For the continuously operated system, solids are pumped directly from the clarifiers into the aerobic digester. In this case, the aerobic digester operates at a fixed liquid level with the overflow going to a solids-liquid separator. Thickened solids are normally recycled back to the digestion tank but when required can also be removed from the system.

Though initially more costly than a batch operated system, much of the manual work involved with aerobic digestion is eliminated.

Another consideration when sizing the aerobic digestion tank is the relationship between the tank geometry desired, the type of aeration equipment being utilized, and the mixing pattern that will develop. An example of an aerobic digester with a mechanical aerator is shown in figure 3–13 and one with diffused aeration equipment is shown in figure 3–14. Figure 3–6 shows the effect of tank geometry on mixing as measured by oxygen transfer rates.

Assume power cost at \$0.03/kwh (\$0.83/mJ), no pacing device on the aeration equipment and that oxygen demand is uniform over 24 hours per day.

Design to handle peak conditions (summer conditions). For 4 Mgal/d (.18 $m^3/s$) was 5,684 lbs oxygen/day (236.8 lbs $O^2$/hr) (107.4 kg/hr). For optimum tank geometry power bill would amount to \$23,225/year. For non-optimum design power bill could get as high as \$38,700/year.

Note that winter conditions use less oxygen, 4,640 lbs/day (193.4 lbs/hr) (87.7 kg/hr). Using a pacing device, savings of \$3,500 to \$5,900/year in power cost could be realized.

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# Thermal Treatment for Sludge Conditioning

## INTRODUCTION

The purpose of this chapter is to consider thermal treatment of sludge as a conditioning process to improve sludge dewaterability by subsequent processes such as vacuum filter, centrifuge or filter press. Thermal conditioning (also often called heat treatment) involves heating sludge, with or without the addition of air or oxygen, to temperatures of 300 to 500°F (150 to 260°C) in a reactor under pressures of 150 to 400 psig (10.5 to 28.1 kgf/cm$^2$) for periods of 15 to 40 minutes. Thermal conditioning causes the release of water and organic material from sludge in the form of a dark brown fluid or "cooking liquor."

Other thermal treatment processes not discussed herein include: (1) pasteurization, which operates at lower temperatures, in the range of 160°F, and (2) wet air oxidation, which operates at higher temperatures and pressures for more complete oxidation of sludge solids.

The EPA Technology Transfer manual on sludge treatment[1] describes thermal conditioning, or heat treatment, as follows:

In heat treatment, temperatures of from 300 to 500°F and pressures of 150 to 400 psig are attained for protracted periods. Significant changes in the nature and composition of wastewater sludges result. The effect of heat treatment has been ideally likened to syneresis, or the breakdown of a gel into water and residual solids. Wastewater sludges are essentially cellular material. These cells contain intracellular gel and extracellular zoogleal slime with equal amounts of carbohydrate and protein. Heat treatment breaks open the cells and releases mainly proteinaceous protoplasm. It also breaks down the protein and zoogleal slime, producing a dark brown liquor consisting of soluble polypeptides, ammonia nitrogen, volatile acids, and carbohydrates. The solid material left behind is mineral matter and cell wall debris.

Dewatering is improved by the solubility and hydrolyzing of the smaller and more highly hydrated sludge particles which then end up in the cooking liquor. While analysis of this liquor from domestic wastewater sludges indicates the breakdown products are mostly organic acids, sugars, polysaccharides, amino acids, ammonia, etc., the exact composition of the liquor is not well defined.

A review of reported analyses of liquor from the heat treatment of sludge gives the range of values shown: $BOD_5$ = 5,000 to 15,000 mg/l, COD = 10,000 to 30,000 mg/l, Ammonia = 500 to 700 mg/l, and Phosphorus as P = 150 to 200 mg/l. About 20 to 30 percent of the COD is not biodegradable in a 30-day period. The volume of cooking liquor from an activated sludge plant with heat treatment amounts to 0.75 to 1.0 percent of the wastewater flow. Based on $BOD_5$ and solids loadings, the liquor can represent 30 to 50 percent of the loading to the aeration system. The pH of cooking liquors is normally in the range of 4 to 5, which necessitates chemical neutralization and/or corrosion resistant equipment.

Figure 4–1 is a flow diagram for a typical heat treatment system. Major components in the system are a heat exchanger and a reaction vessel. Heat treatment

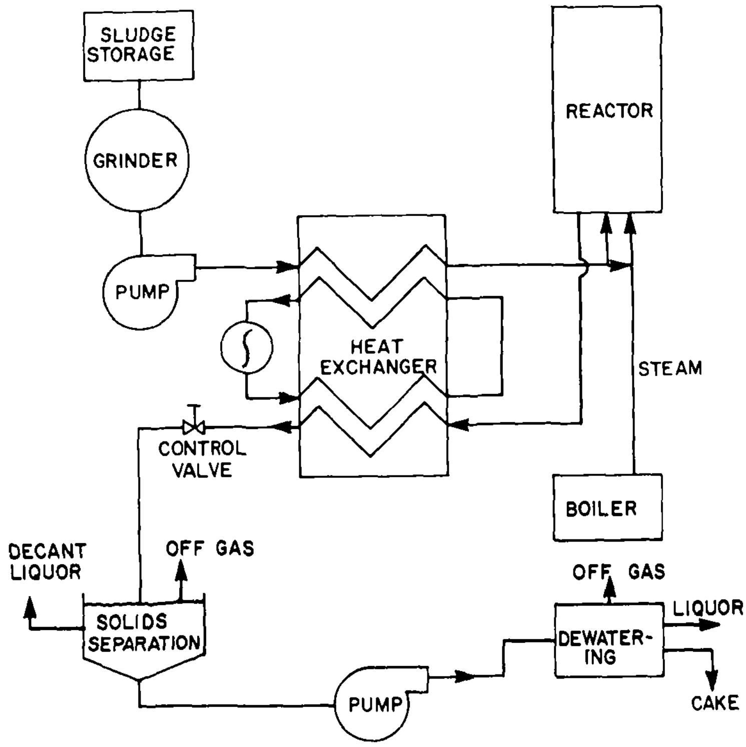



Figure 4–1.—Typical heat treatment system.

may be used to condition raw or digested sludges and thus location of the system in the overall treatment train may vary. If a treatment plant employs anaerobic digestion, heat treatment is more commonly used to condition the digested sludge. Heat treatment before anaerobic digestion to improve degradability and energy production was pilot tested by LA/OMA in Los Angeles.[2,3] Heat treatment may be used in conjunction with incineration in a system that recycles waste heat to minimize energy requirements. These variations in the use of heat treatment in sludge management systems are illustrated in figure 4–2.

The effect of heat treatment on the chemical composition of sludge was investigated by Sommers and Curtis.[4] Heat treated sludges from plants in Speedway and Terre Haute, Indiana were tested to obtain information on the forms of nitrogen, phosphorus, copper, zinc, nickel, cadmium and lead. In general, heat treatment produced greater than 50 percent reductions in total nitrogen with essentially no change or a slight increase in phosphorus and metals concentrations.

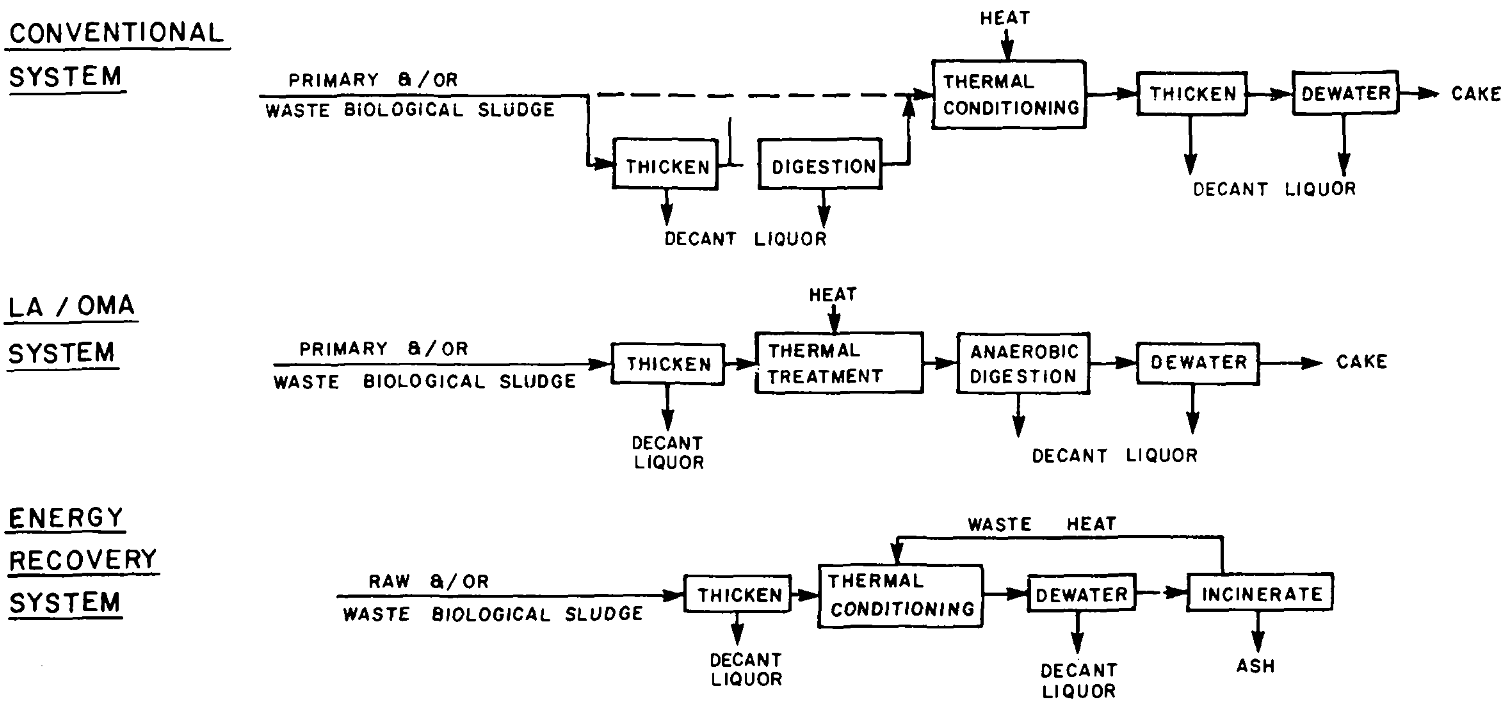



Figure 4–2.—Heat treatment in sludge management systems.

## PROCESS DESCRIPTION

Equipment for thermal conditioning of sludge is manufactured and supplied in the United States by Envirotech BSP (Porteous System), Zimpro (wet oxidation), Zurn, and Nichols. Almost all of the equipment for thermal conditioning of sludge in the United States has been supplied by Zimpro or Envirotech. Mayer and Knopp[4] reported in January 1977, that 70 thermal conditioning plants were operating in the United States and Canada and 43 others were under construction.

| *Type of plant* | *Number of installations* |
|---|---|
| With air addition | |
| Operating | 45 |
| Under construction | 35 |
| Without air addition | |
| Operating | 25 |
| Under construction | 8 |

A partial list of thermal conditioning installations is shown in table 4–1.

### Zimpro Process

The Zimpro system is similar to the process illustrated in figure 4–1 except that air is also added to the reactor. Basic features of the Zimpro process are (1) air addition to the reactor for oxidation, improvement of heat exchange characteristics and reduction of fuel requirements, and (2) use of sludge-to-sludge heat exchanger. Some of the equipment used in this process is shown in figures 4–3 and 4–4.

In the continuous process, the sludge is passed

Table 4–1.—Size and status of largest thermal conditioning installations

| Location | Status | Number of units | Unit capacity (gal/min) |
|---|---|---|---|
| Toronto, Ontario (Ashbridges Bay) | UC[a] | 7 | 250 |
| Cleveland, Ohio (Southerly) | UC | 5 | 280 |
| Louisville, Ky. | Operating (1976) | 4 | 250 |
| Cincinnati, Ohio (Mill Creek) | Operating | 4 | 280 |
| Flint, Mich. | Operating | 3 | 250 |
| Green Bay, Wis. | Operating (1975) | 4 | 150 |
| Columbus, Ohio (Southerly) | Operating (1976) | 3 | 200 |
| Suffolk Co., N.Y. | UC | 2 | 205 |
| Toronto, Ontario (Lakeview) | Operating (1975) | 3 | 125 |
| Springfield, Mass. | UC | 2 | 200 |
| Kalamazoo, Mich. | Operating (1971) | 3 | 125 |
| Columbus, Ohio | Operating (1972) | 1 | 200 |
| Toronto, Ontario (Highland Creek) | UC | 3 | 125 |
| Chesapeake-Elizabeth, Va. | UC | 1 | 150 |
| Hopewell, Va. | UC | 3 | 150 |
| York, Pa. | UC | 2 | 125 |
| Billings, Mont. | UC | 2 | 100 |
| Escondido, Calif. | UC | 1 | 100 |

[a]Under construction.

Figure 4–3.—Reactor (left), heat exchangers (center), waste heat recovery boiler (right).

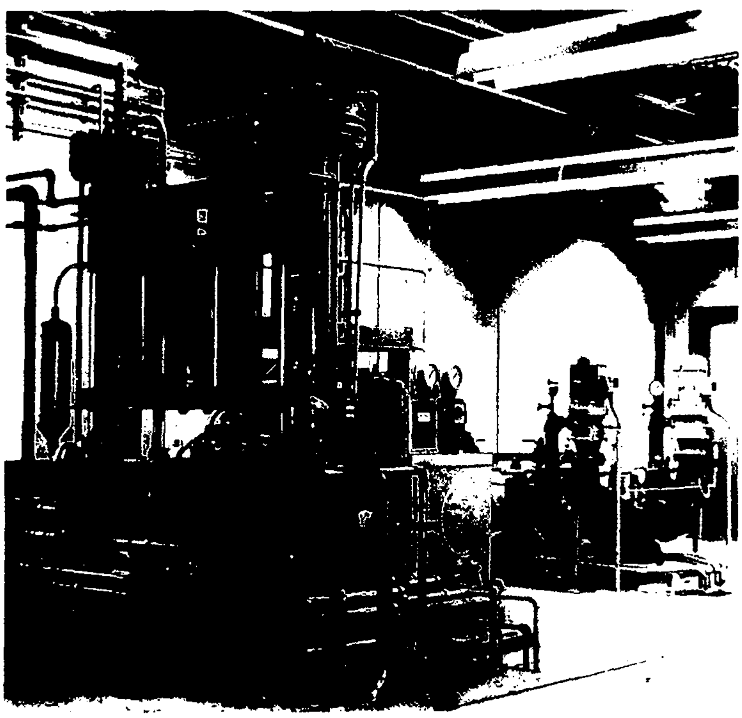


Figure 4–4.—Pump (left), grinder (right).

through a grinder which reduces the size of sludge particles to about one-fourth inch. Sludge and air are then pumped into the system and the mixture is passed through heat exchangers and brought to the initial reaction temperature. As oxidation takes place in the reactor, the temperature increases. The oxidized products leaving the reactor are cooled in the heat exchanger by the entering cold sludge and air. The gases are separated from the liquid carrying the residual oxidized solids, usually in a decant tank, and released through an odor control unit. The oxidized liquid and remaining suspended solids are separated in a decant tank. The decant tank underflow may be further dewatered by several methods; the overflow cooking liquor is recycled to the main plant or treated by a separate system such as activated sludge, rotating biological disk or anaerobic filter.

## Envirotech BSP Process

This system was formerly called the Porteus process. The Porteous process was purchased by Envirotech and various changes have been made in the system. The basic system components and operation of the BSP system are similar to the Zimpro process as illustrated in figure 4–1. One basic difference is that air is not injected into the reactor in the BSP system. The BSP systems also normally employ a water-to-sludge heat exchanger.

## Other Processes

The Nichols heat treatment system was previously marketed as the Dorr-Oliver Farrer system. The Nichols process is used at a plant serving York, Pa., and there are five installations of the Farrer system in the United States: San Bernardino, Calif.; Elkhart, Ind.; Port Huron, Mich.; Glouster, N. J.; Norwalk, Conn. There is a Zurn heat treatment system in Mentor, Ohio which serves an area of Lake County, Ohio.

## Thermal Treatment Process Sidestreams

There are both liquid and gas byproducts from any thermal conditioning system. These sidestreams must be considered in planning for an accurate comparison with other processes and in design for a properly operating system.

### *Gas Sidestreams*

There are four principal sources of odor resulting from thermal sludge treatment: (1) vapors from treated sludge storage (decant tank or thickener), (2) mechanical dewatering system exhaust, (3) exhausted air from working atmosphere in filter and loading hopper areas, and (4) vapors from strong liquor pretreatment devices. The odorous gases produced are simple, low molecular weight, volatile organic substances, consisting of aldehydes, ketones, various sulphurous compounds, and organic acids. The odor level source associated with thermal sludge conditioning is dependent to a high degree on the total hydrocarbon content. The odor level and hydrocarbon content are highest in off-gases from the heat treated sludge solids separation units, i.e., decant tank or thickener and mechanical dewatering systems.

Off-gases are best controlled by use of incineration, adsorption, or scrubbing (or some combination of these processes).

Water scrubbing plus incineration.—For high hydrocarbon airstreams, the highest degree of odor control can be obtained by water scrubbing followed by incineration. The scrubbing portion of this system consists of a packed bed unit which uses plant effluent water at rates of 20 to 30 gpm (1.3 to 1.9 l/s) per 1,000 ft$^3$/m (472.0 l/s). The incineration portion of this system can be either direct flame incineration at 1,500°F (815°C) or catalytic incineration at 800°F (427°C). The oxidation catalysts that are commonly used in catalytic incineration are supported platinum or palladium materials.

Water scrubbing plus adsorption.—In scrubbing methods, the odorous substances are removed by solubilization, condensation, or chemical reaction with the scrubbing medium. Scrubbing media that are commonly used for odor control are potassium permanganate, sodium hydroxide, or sodium hypochlorite. Two to four pounds of potassium permanganate are required per pound of hydrocarbon removed. In the adsorption method, substances are removed from the odorous gas stream by adsorption on activated carbon or silica gel. The activated carbon or silica gel must be capable of regeneration, usually by steaming. High hydrocarbon sources can be treated in an odor control system composed of a water scrubber followed by an activated carbon adsorption unit. The water scrubber is the same as that described above. The carbon adsorption unit is a multiple bed adsorber that is sized to minimize the required number of steam regenerations. Normally, the carbon bed would be sized so that only one steam regeneration per day would be required. Treating a 1,000 ft$^3$/min (472.0 l/s) gas stream would require a dual bed carbon system containing 1,800 pounds (816 kg) of carbon per bed. This sizing would permit an adsorption cycle of 24 hours. After a 24-hour adsorption time, the second carbon bed would be placed in the adsorption cycle and the spent bed would be steam regenerated. The regeneration cycle requires low pressure steam at a maximum of 50 psig (3.5 kgf/cm$^2$) for a period of one hour. The steam and desorbed organic compounds from the bed are condensed and collected. The aqueous condensate is returned to the head of the treatment plant and the liquid organic phase is incinerated.

Multiple scrubbers.—A third option for treating high hydrocarbon sources is a multiple scrubber system. The multiple scrubber system would contain at least two and preferably three scrubbing stages. In all cases, the first scrubbing stage of the system should be water scrubbing using plant effluent at a rate of about 27 gal/min (1.7 l/s) per 1,000 ft$^3$/min (472.0 l/s). The second and third stages should be chemical scrubbing using a combination of scrubbing media selected from 5 percent sodium hydroxide, 3 percent sodium hypochlorite, and 3 percent potassium permanganate. The potassium permanganate solution effects the highest degree of hydrocarbon reduction and, hence, the highest odor reduction. One of the most effective multiple scrubber systems consists of three stages utilizing plant effluent, 5 percent sodium hydroxide and 3 percent potassium permanganate.

## Liquid Sidestreams

The liquid (cooking liquor) containing materials solubilized during heat treatment of sludge may be separated from the solids (1) during storage in decant tank, thickener, or lagoon, and (2) in the dewatering step using a vacuum filter, centrifuge, filter press, sand drying bed or other method.

The following tabulation shows some of the substances present in thermal treatment liquor and the general ranges of concentration.

| *Constituent* | *Concentration range mg/l (except color)* |
|---|---|
| Suspended solids | 100–20,000 |
| Chemical oxygen demand | 10,000–30,000 |
| Biochemical oxygen demand | 5,000–15,000 |
| Ammonia nitrogen | 500– 700 |
| Phosphorus | 150– 200 |
| Color, units | 1,000– 6,000 |

The composition of thermal treatment liquor varies widely depending upon sludge composition and reactor operating conditions. For a given reactor temperature, increasing the reactor detention time will generally increase the concentration of soluble organic material in the cooking liquor. Heat treatment can normally be expected to solubilize from 40 to 70 percent of the sludge biomass. As much as 60 to 70 percent of the suspended solids in waste activated sludge were solubilized in heat treatment pilot tests in Los Angeles.[6]

The character of the cooking liquor is somewhat uncertain and the subject of some debate. The EPA Sludge Manual[1] states: "About 20 to 30 percent of the COD is not biodegradable in a 30-day period." Based on pilot scale heat treatment studies of mixed primary and waste activated sludge, Erickson and Knopp[7] estimated that the soluble nonbiodegradable COD was 7 percent of the total cooking liquor COD. Laboratory tests by Stack, et al.,[8] indicated that about 40 percent of organics in the cooking liquor from heat treatment of waste activated sludge were resistant to biological oxidation.

The EPA Sludge Manual further states: "Based on $BOD_5$ and solids loadings, the liquor can represent 30 to 50 percent of the loading to the aeration system." Boyle and Gruenwald[9] reported that the heat treatment recycle liquor BOD contributed approximately 21 percent of the BOD entering the Colorado Springs, Colorado plant. Studies by Haug, et al.,[6] indicated that recycle of cooking liquor in the Hyperion plant at Los Angeles would increase the oxygen demand on the aeration system by about 30 percent.

Thermal treatment liquor can be treated by recycle to the main treatment plant or by separate treatment systems such as activated sludge, rotating biological disks or anaerobic filters.

Recycle to main plant.—Thermal treatment liquor often is recycled through the main treatment plant, being added to the raw sewage or primary effluent. This places an additional load upon the system principally in the form of oxygen demand, suspended solids and color. In most cases the color and COD of the final effluent will increase. The effects of recycle can be mitigated to some extent by storing thermal treatment liquor and returning it to the treatment plant at a uniform rate or during off-peak hours.

Separate treatment and disposal.—Another method for handling liquor is to treat the sidestreams separately with no return of any liquor to the main treatment plant. Sometimes digester supernatant and waste activated sludge are combined with the thermal treatment liquor for separate processing; one example of this method is the installation at Indio, Calif. where aerated lagoons with long retention provide satisfactory results. Lagoon effluent is blended with plant effluent for discharge.

Separate treatment prior to recycle.—In order to reduce the load on the main treatment plant and maintain final effluent quality, cooking liquor is often treated separately prior to recycle to the main plant. Plain aeration, extended aeration, and activated sludge treatment have been used for pretreatment of cooking liquors. BOD reductions by conventional activated sludge pretreatment of liquors have been reported as high as 90 percent. It may be necessary to collect and deodorize aeration basin off-gases.

## THERMAL CONDITIONING COSTS

Thermal conditioning of sludge affects the cost of other treatment plant processes, decreasing some and increasing others. Total cost includes direct capital, operating, and maintenance costs for sludge handling plus or minus the indirect net cost effect of sludge handling on other treatment plant processes. Added costs resulting from heat treatment include: (1) cooking liquor treatment, and (2) treatment of odorous off-gases. Potential cost savings include reduction in subsequent sludge treatment and disposal costs because of improved sludge dewatering.

An EPA[10] report presents detailed cost estimates for thermal conditioning and sidestream treatment. Costs were based on data from several sources including operating plants, published literature, manufacturers data and engineering estimates. The following cost information was developed for thermal conditioning systems (does not include costs for sidestream treatment):

1. Capital costs for thermal systems vary from about $50,000 per gal/min ($790,000 per l/s) of thermal treatment system capacity for a 10 gal/min (.6 l/s) system to $10,000 per gal/min ($159,000 per l/s) for a 200 gal/min (12.6 l/s) system.
2. Typical fuel requirements are 900 to 1,000 Btu per gallon (249 to 277 kJ/l) for systems that do not employ air addition and 300 to 600 Btu per gallon (83 to 166 kJ/l) with air addition.
3. Average electrical energy consumption averaged 22 kWh per 1,000 gallons (209 J/l) for plants with air addition and 10 kWh per 1,000 gallons (95 J/l) without air addition.
4. Operation and maintenance labor constitutes a significant fraction of overall costs, ranging from 6,000 hours per year for a 10 gal/min (.6 l/s) system to 20,000 hours per year for a 200 gal/min (12.6 l/s) system.
5. Costs for materials and supplies range from $5,000 per year for a 10 gal/min (.6 l/s) system to $20,000 per year for a 200 gal/min (12.6 l/s) system.

The following cost information is related to indirect costs for treating sidestreams:

1. Increased capital costs primarily result from the need to increase aeration tank volume and air supply capabilities.
2. Increased energy is required for aeration capacity required to treat the recycled liquor.
3. Increased labor is required for maintaining and operating the additional aeration capacity and related settling and pumping systems.

Costs for treating the off-gas from the thermal treatment system typically constitutes 5 to 10 percent of the total cost for thermal treatment. Carbon adsorption is the most costly technique for odor control. Incineration is most economical in smaller plants and chemical scrubbing in larger plants.

Based on unit costs of $7 per hour for labor, $0.03 per kWh for electricity, and $2.80 per million Btu and amortization of capital costs over 20 years at 7 percent interest, the following typical costs for thermal conditioning were determined (all costs are dollars per ton of dry solids processed):

| | *Construction costs* | | | *O. & M. cost* | | | |
|---|---|---|---|---|---|---|---|
| *Sludge ton/day* | *Direct* | *Indirect* | *Total* | *Direct* | *Indirect* | *Total* | *Total* |
| 1 ....... | 97.53 | 4.11 | 101.64 | 150.14 | 4.93 | 155.07 | 256.71 |
| 5 ....... | 30.79 | 3.18 | 33.97 | 46.46 | 3.67 | 50.13 | 84.10 |
| 10 ....... | 21.45 | 2.93 | 24.38 | 32.52 | 3.50 | 36.02 | 60.40 |
| 50 ....... | 12.20 | 1.83 | 14.03 | 19.10 | 2.99 | 22.09 | 36.12 |
| 100 ....... | 10.96 | 1.98 | 12.94 | 16.58 | 2.87 | 19.45 | 32.39 |

The March 1975 national average construction costs for thermal conditioning are shown on figure 4–5. These costs include feed pumps; grinders; heat exchangers; reactors; boilers; gas separators; air compressors where applicable; decanting tanks; standard odor control systems; and piping, controls, wiring and installation services usually furnished by the equipment or system manufacturer. Not included in the basic thermal treatment costs are buildings: footings; piping; electrical work and utilities not supplied by the equipment manufacturer;

sludge storage and thickening prior to thermal treatment; sludge dewatering, incineration or diposal; land; and engineering fees. In escalating costs for later dates, it should be considered that the escalation determined from the EPA-STP index may not adequately reflect the increased costs for high temperature, equipment-dominated processes such as thermal treatment.

A second curve (curve B) is shown on figure 4–5 and includes the costs for typical building, foundation and utility needs for thermal treatment systems. The building costs represent single-story, concrete or masonry construction with built-up roofing, insulation and heat and vent systems, and assume that reactors and decant tank will be located outside of the building. The costs also include piping and wiring within the building, foundations for internal and external equipment, and limited amount of site work. Building sizes provide for easy access to equipment and control room. For larger installations, where multiple units are anticipated, space for some standby equipment is included. Typical building sizes range from 1,500 square feet (139 $m^2$) for a 10 gal/min (.6 l/s) plant to 5,250 square feet (488 $m^2$) for a 200 gal/min (12.6 l/s) plant. The construction cost of the building was estimated to be \$36/$ft^2$ (\$387/$m^2$).

The curves show a rapid rise in unit construction costs for plants smaller than about 20 gal/min (1.3 l/s). The minimum direct cost of a thermal treatment plant is estimated to be approximately \$350,000 regardless of plant size. For plants above about 150 gal/min (9.5 l/s) the increased use of multiple treatment units and of standby units results in a lower limit for unit cost per gal/min of capacity. This lower limit appears to be in the range of \$9,000 to \$12,000 per gal/min (\$143,000 to \$190,000 per l/s). Data for these larger plants are sparse, however, and some plants reported lower unit costs.

The annual fuel requirements based on 8,000 hours of operation are shown in figure 4–6. Fuel is used chiefly as a source of heat to produce steam. The amount of fuel used is influenced by the reaction temperature, efficiencies of the boiler and heat exchange systems, insulation or heat losses from the system and the degree of heat-producing oxidation which takes place in the reactor. Some reduction in the unit heat requirement for increase in plant size is reported. This is believed to result from more uniform and constant operation of the system, greater heat transfer and insulation efficiencies and possibly a greater amount of oxidation in the larger units. Plants adding air to heat exchangers and reactors experiencing some oxidation have lower fuel requirements.

Typical fuel requirements averaged 900 to 1,000 Btu per gallon (249 to 277 kJ/l) for plants not practicing air addition and 300 to 600 Btu per gallon (83 to 166 kJ/l) depending on the degree of oxidation obtained, for plants practicing air addition. Curves in this paper are based on fuel requirements of 900 Btu per gallon (249 kJ/l) for thermal conditioning without air and 500 Btu

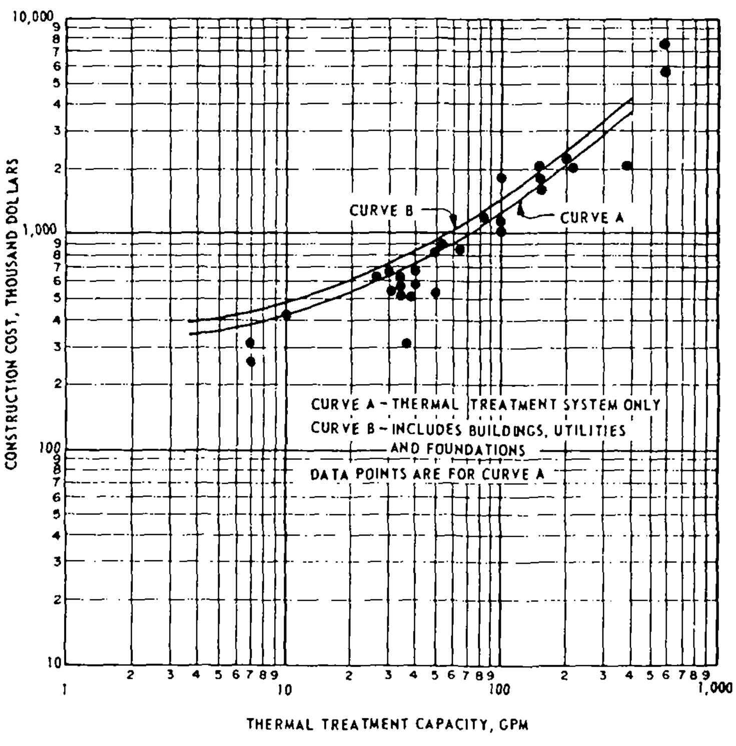



Figure 4–5.—Direct construction costs for thermal conditioning.

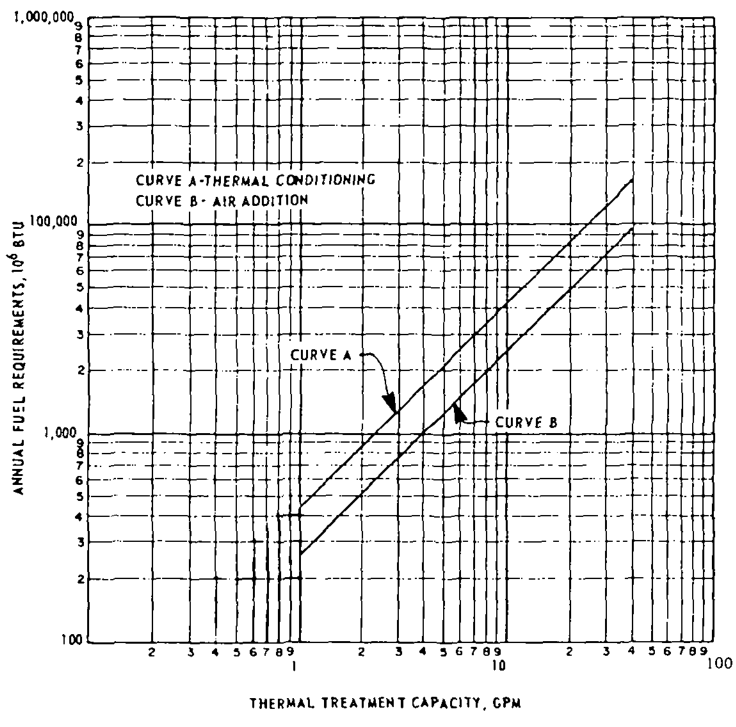



Figure 4–6.—Annual direct fuel requirements for thermal conditioning.

per gallon (139 kJ/l), corresponding to about five percent oxidation, plants with air addition. These fuel requirements do not include allowances for treatment of off-gas.

Annual electrical energy usages for the two types of plants (with and without air addition) are shown in figure 4–5. A separate curve is included on figure 4–7 for estimating the energy requirements for building needs. Electrical energy requirements are determined by sizes and efficiencies of machinery such as sludge and boiler water pumps, grinders, thickeners and, in plants where air addition is practiced, air compressors. Electrical energy is also required for lighting and other building uses. Average unit energy requirements are 22 kWh per 1,000 gallons (209 J/l) for plants practicing air addition and 10 kWh per 1,000 gallons (95 J/l) for plants without air addition.

Operation and maintenance labor requirements are shown in figure 4–8. In this paper operation comprises time spent collecting and logging data on the process, controlling and adjusting the various systems and components, and laboratory work. The functions covered by maintenance include cleaning and repairing process components, general upkeep of the process area, checking and repairing of controls and instrumentation, and performing preventative maintenance. Maintenance in figure 4–8 does not include major overhauls which will be required periodically. In some plants these operation and maintenance functions may vary or may overlap.

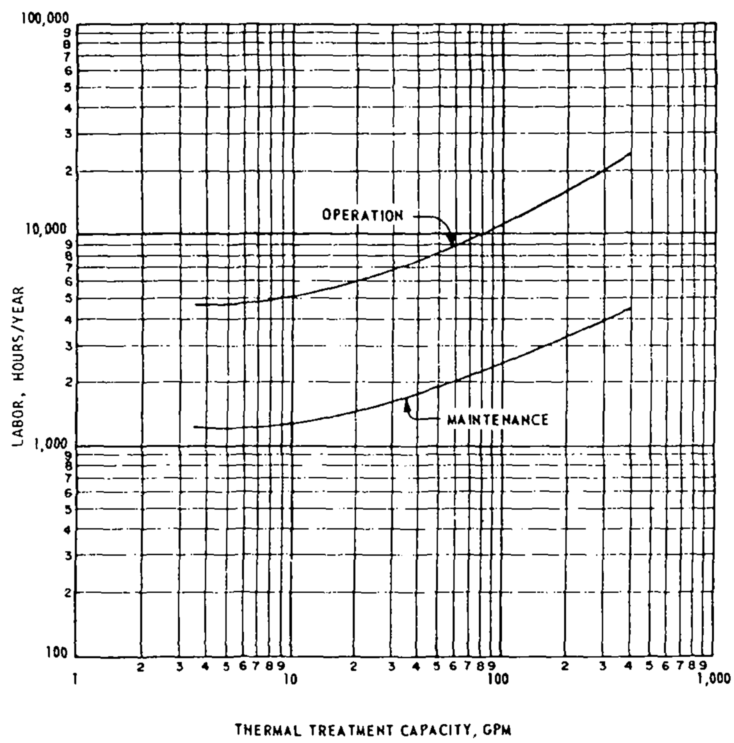



Figure 4–8.—Operating and maintenance labor requirements for thermal conditioning.

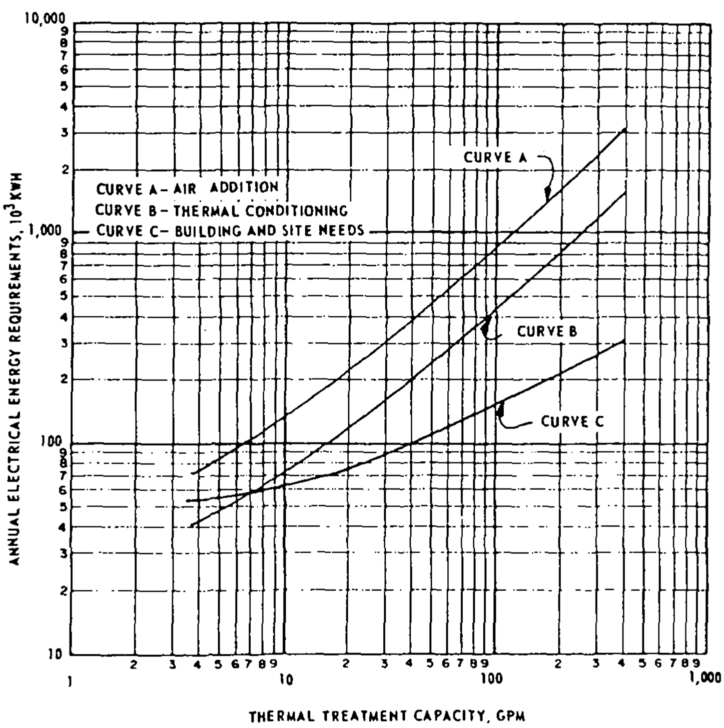



Figure 4–7.—Annual direct electrical energy requirements for thermal conditioning.

In general, maintenance labor is approximately one-fourth of operating labor, ranging from the equivalent of one maintenance man for one shift at a 50 gal/min (3.2 l/s) plant to one and one-half men for one shift at a 200 gal/min (12.6 l/s) plant. The amount of maintenance required depends greatly on the design and operation of the plant, particularly on equipment and materials used for construction. It is also dependent on the skill and knowledge of the maintenance personnel and the design of, and adherence to, a preventative maintenance program.

Annual costs for materials and supplies are shown in figure 4–9. Curve A shows the normal annual costs for materials and supplies required to operate and maintain the thermal conditioning system. These costs are plotted against thermal treatment plant capacity and include materials and parts such as seals, packing, coatings, lamps, bearings, grinder blades, and other items used in scheduled and normal maintenance. They also include operating supplies such as lubricants, cleaning chemicals, boiler feed water, and water treating chemicals. These costs vary from about $5,000 per year for a 10 gal/min (.6 l/s) plant to approximately $20,000 per year for a 200 gal/min (12.6 l/s) plant.

Besides normal, periodic maintenance required for a plant shown by curve A, additional costs for major overhaul work are incurred. This work includes such items as motor rewinding; major overhauls of pumps and com-

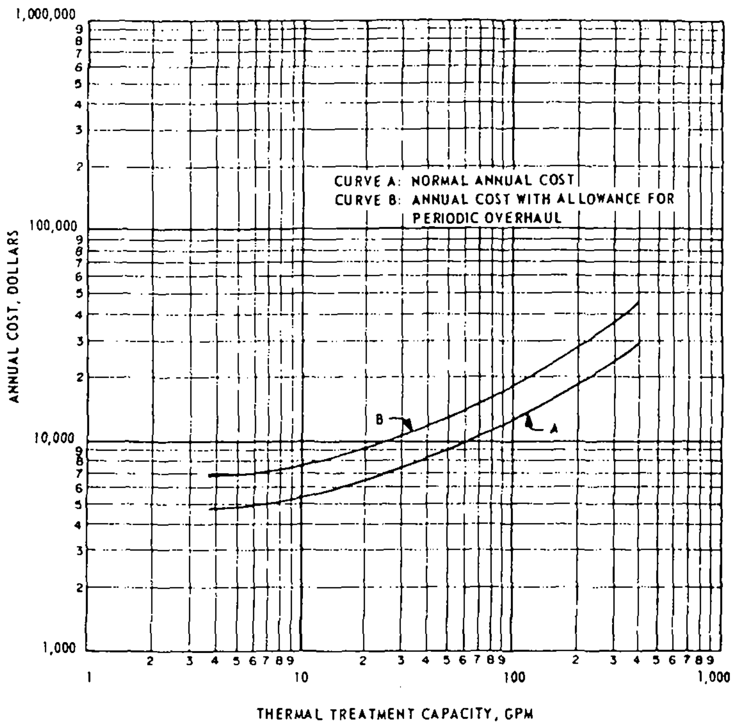



Figure 4–9.—Materials and supplies for thermal conditioning.

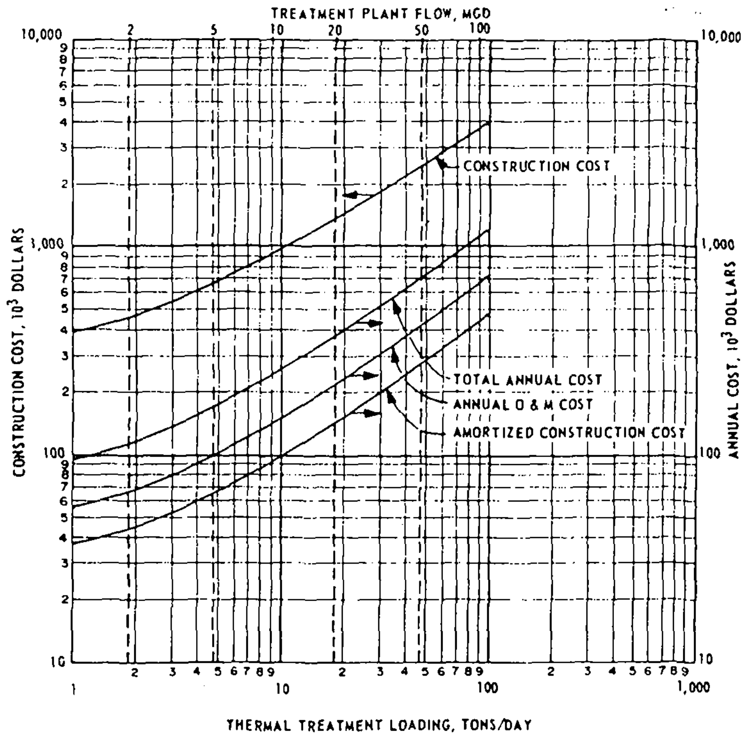



Figure 4–10.—Direct and indirect costs for thermal conditioning.

pressors; major non-routine rehabilitation or replacement of heat exchanger tubing piping and controls; and refitting of boilers. This type of work is required at an average interval of about 6 to 7 years, depending on the conditions at a particular plant. Because labor for this type of major work is often contracted, labor costs are treated as part of the overhaul and included in its cost under this section. Curve B shows the combination of these costs with those included under curve A to give the total annual cost for the materials and supplies. The inclusion of major overhaul work increases the annual materials cost by about 45 percent over that required for routine and preventative maintenance materials.

There was considerable variation among the costs for materials in seemingly similar plants and it appeared that three factors tended to govern the costs.

1. Preventative maintenance program. In plants where a good program was practiced, overall costs for parts supplies generally were lower. Where maintenance was neglected, more major failures were found to occur with a need for greater expenditure for parts.
2. Design of the plant and selection of materials of construction. If a higher grade of materials and equipment were selected for initial construction and if the plant were designed with ease of maintenance in mind, less maintenance and better maintenance were found and hence less need for replacement was noted.
3. Quality of the water supply. In areas with high hardness and high mineral contents in their water supplies, more scaling and corrosion were noted in equipment, particularly in heat exchangers. Scaling, along with the increased amount of cleaning required, resulted in both an increase in replacement parts for boilers and heat exchangers and an increased amount of chemicals for boiler water treatment and heat exchanger cleaning.

Total costs for thermal conditioning systems, with air addition, including costs for treatment of cooking liquor and odorous gas sidestreams are shown in figure 4–10. Costs in figure 4–10 are based on the following:

1. Cooking liquor treated in the main plant by increasing the size of activated sludge system.
2. Capital costs include an allowance for engineering, legal and administrative and interest during construction and amortized over 20 years at 7 percent interest.
3. Electrical energy cost = \$0.03/kWh (\$0.83/mJ).
4. Fuel cost = \$2.80/million Btu (\$2.65/GJ).
5. Labor cost = \$7.00/hour.

Using the above criteria, total costs for thermal conditioning range from \$257/ton (\$283/Mg) in a 1 ton/day (0.9 Mg/day) capacity plant to \$32/ton (\$35/Mg) in a 100 ton/day (91 Mg/day) plant.

## DESIGN EXAMPLE

The design example considered herein is a 4 Mgal/d standard activated sludge plant with the following sludge characteristics:

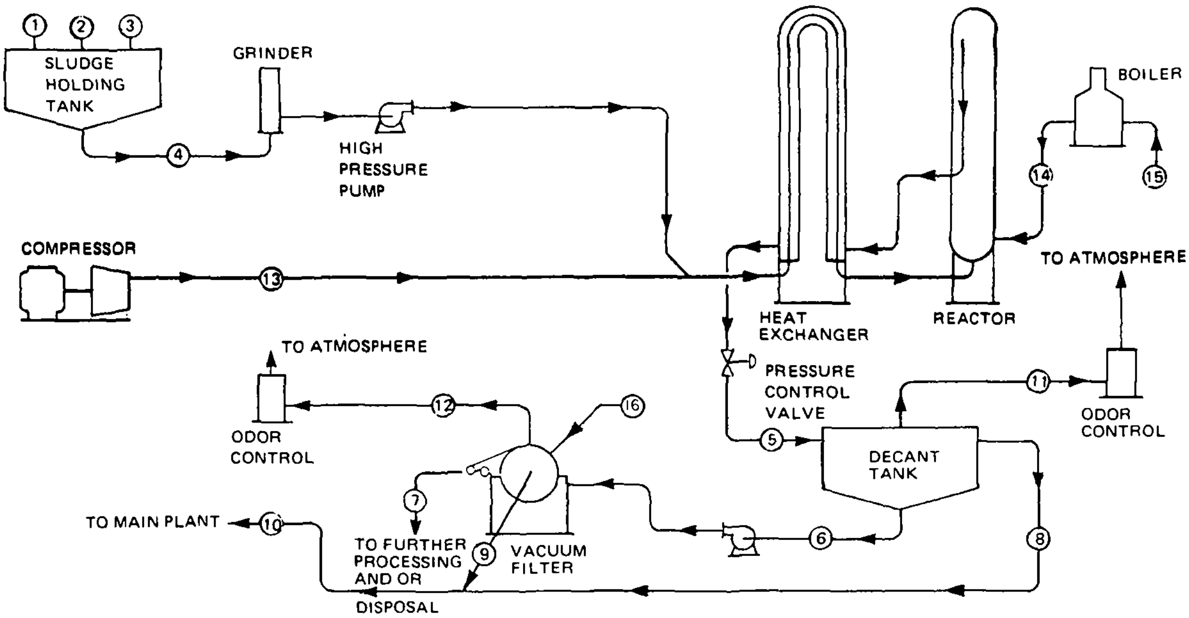
| *Sludge type* | *Total solids (lb)* | *Volatile solids (lb)* | *Flow (gal/min)* | *Flow (Mgal/d)* |
|---|---|---|---|---|
| Primary | 5,200 | 3,120 | 5.4 | 0.008 |
| Secondary | 4,000 | 3,200 | 8.3 | 0.012 |
| Total | 9,200 | 6,320 | 13.7 | 0.020 |

These sludge quantities were determined with the following assumptions:

1. Raw wastewater suspended solids = 240 mg/l; BOD = 200 mg/l.
2. Suspended solids removal = 65 percent in primary treatment and 90 percent overall; BOD removal = 30 percent in primary treatment and 90 percent overall.
3. One-half pound activated sludge produced per pound BOD removed.
4. Primary sludge is 4 percent solids and is gravity thickened to 8 percent solids.
5. Waste activated sludge is 1 percent solids and is thickened to 4 percent solids.

A process and materials flow diagram is shown in figure 4–11 for a thermal conditioning system of primary and secondary sludge. The example system utilizes air addition and assumes that the recycle liquor will be treated in the main activated sludge plant. Other features of this system include the following:

1. One thermal conditioning reactor required.
   Flow = 20 gpm (1.3 l/s)
   Operating pressure = 350 psig (24.6 kgf/cm$^2$)
   Operating temperature = 370° F (225° C)
   Operating schedule: 24 hours/day, 7 days/week
   Installed horsepower = 85 (63.5 kW)




| LOCATION | Mgal/d | gmp | Ton/day | Total Solids lb/day | Percent Solids | BOD$_5$ lb/day | BOD$_5$ mg/l |
|---|---|---|---|---|---|---|---|
| 1. Primary Sludge | 0.008 | 5.5 | 32 | 5,230 | 8.0 | – | – |
| 2. Secondary Sludge | .012 | 8.4 | 50 | 4,040 | 4.0 | – | – |
| 3. Recycled Sludge | .002 | 1.4 | 11 | 830 | 3.6 | – | – |
| 4. Total Sludge | .022 | 15.3 | 93 | 10,100 | 5.4 | – | – |
| 5. Conditioned Sludge | .023 | 16.0 | 98 | 9,760 | 5.1 | – | – |
| 6. Decant Underflow | .009 | 6.3 | 36 | 8,015 | 11.1 | – | – |
| 7. Vacuum Filter Cake | – | – | 10 | 7,200 | 36.0 | – | – |
| 8. Decant Supernatant | .015 | 10.4 | 61 | 1,730 | – | 875 | 7,000 |
| 9. Vacuum Filter Filtrate | .013 | 9.0 | 56 | 840 | – | 370 | 3,400 |
| 10. Total Liquid Recycle | .028 | 19.4 | 117 | 2,570 | 1.1 | 1245 | 5,300 |
| 11. Decant Tank Exhaust–81 scfm | | | | | | | |
| 12. Vacuum Filter Exhaust–2400 scfm | | | | | | | |
| 13. Air to Reactor–32 scfm | | | | | | | |
| 14. Steam to Reactor–8,000 lb/day | | | | | | | |
| 15. Boiler Feed Water–0.001 Mgal/d (0.7 gpm) | | | | | | | |
| 16. Vacuum Filter Wash Water–0.007 mgd (5 gpm) | | | | | | | |

Figure 4–11.—Thermal conditioning example 4 Mgal/d activated sludge plant.

Building area required = 1,115 square feet (103.6 $m^2$)

2. One decant tank required.
   Design loading = 50 lb/sq ft/day (244 $kg/m^2/day$)
   Diameter = 15 feet (4.57 m)
   Side water depth = 10 feet (305 m)
3. Scrubber-afterburner system to treat 81 scfm (38.2 l/s) odorous gas from decant tank.
   Installed horsepower = 3 (2.2 kW)
   Building area required = 32 square feet (3.0 $m^2$)
4. Multi-stage scrubber to treat 2,400 scfm (1130 l/s) odorous gas from vacuum filter.
   Installed horsepower = 13 (9.7 kW)
   Building area required = 144 square feet (13.4 $m^2$)

In this example, the assumed BOD loading without thermal conditioning is 6,670 pounds (3025 kg) per day in the raw wastewater and 4,670 pounds (2118 kg) per day to the aeration basins. The BOD in the decant tank supernatant and the vacuum filter filtrate are estimated to increase the main treatment plant loading as follows:

| | *Decant tank supernatant* | *Vacuum filter filtrate* | *Total recycle flow* |
|---|---|---|---|
| $BOD_5$, lb/day | 875 | 370 | 1,245 |
| Percent $BOD_5$ in raw wastewater | 13.1 | 5.6 | 18.7 |
| Percent $BOD_5$ to aeration basins | 18.7 | 7.9 | 26.7 |

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# Thickening of Sludge

## INTRODUCTION

Sludge thickening is defined as increasing the total solids concentration of a dilute sludge from its initial value to some higher value, up to a limit of about 10–12 percent total solids. Thickening is contrasted with "dewatering" which increases the total solids concentration to the range of 15–30 percent. Thickening operations are intended to reduce the volume of sludge to be further processed and normally constitute an intermediate step preceding dewatering or stabilization.

The unit processes most commonly associated with wastewater sludge thickening are gravity thickening, dissolved air flotation, and centrifugation. Some of the heavier sludges, such as raw primary and combinations of raw primary and some biological sludges, may be readily thickened with gravity thickeners. Other, more flocculent sludges, such as those from activated sludge processes, may require more elaborate methods. The most frequent applications of the common processes are summarized in table 5–1.

The selection and design of a sludge thickening system is dependent upon many factors including the sludge characteristics, sludge processing following thickening, and the type and size of wastewater treatment facility. Each thickening situation will be somewhat different. Applications other than those shown in table 5–1 are possible and, in some cases, may provide the desired results.

This paper will discuss in detail the processes of gravity thickening, dissolved air flotation, and centrifugation. Other newer methods will also be mentioned. First, sludge characteristics and sludge handling methods will be discussed. This will be followed by a discussion of the thickening processes, performance data, and recommended design standards. This material will then be used in a design example which will illustrate the general approach necessary in thickening alternative evaluation and selection. Bench scale or pilot studies are frequently required for determining applicability of, and/or design parameters for, the various thickening processes. Examples of these will be presented with the design example. Additionally, equipment capital, operation, and maintenance cost data will necessarily be presented to aid in screening the alternatives. As the example is developed, the methodology for determining the most reliable and cost effective process for a given sludge will be shown.

Table 5–1.—Frequent applications of thickening processes

| Process description | Sludge applications |
|---|---|
| Gravity thickening | Primary sludge<br>Combined primary and secondary sludges |
| Dissolved air flotation | Secondary sludges |
| Centrifugation | Secondary sludges |

## SLUDGE CHARACTERISTICS AND HANDLING

Separation of solid matter from wastewater in a settling tank results in a clarified tank effluent and a watery mass of solids known as "sludge." Many different sludge types and variations in sludge concentration are encountered in wastewater treatment. The characteristics of a sludge prior to thickening will generally depend upon the type of wastewater treated, the sludge origin (which particular wastewater treatment process), the degree of chemical addition for improved settling or phosphorus removal, and the sludge age. Additionally, the sludge produced by a specific settling tank will also depend somewhat upon the design and operation of the unit. Typical "as removed" sludge concentrations are presented in table 5–2.

Table 5–2.—Typical sludge characteristics "as removed" from treatment processes

| Sludge type | Range percent solids | Typical percent solids |
|---|---|---|
| Primary (PRI) | 2–7 | 4 |
| Waste activated (WAS) | 0.5–1.5 | 1 |
| Extended aeration (EA) | 1–3 | 2 |
| Trickling filter (TF) | 1–4 | 2 |
| Rotating biological disc (RBD) | 1–3.5 | 2 |
| Combinations: | | |
| PRI + WAS | 2.5–4 | 3 |
| PRI + TF | 2–6 | 3.5 |
| PRI + RBD | 2–6 | 3.5 |
| WAS + TF | 0.5–2.5 | 1.5 |

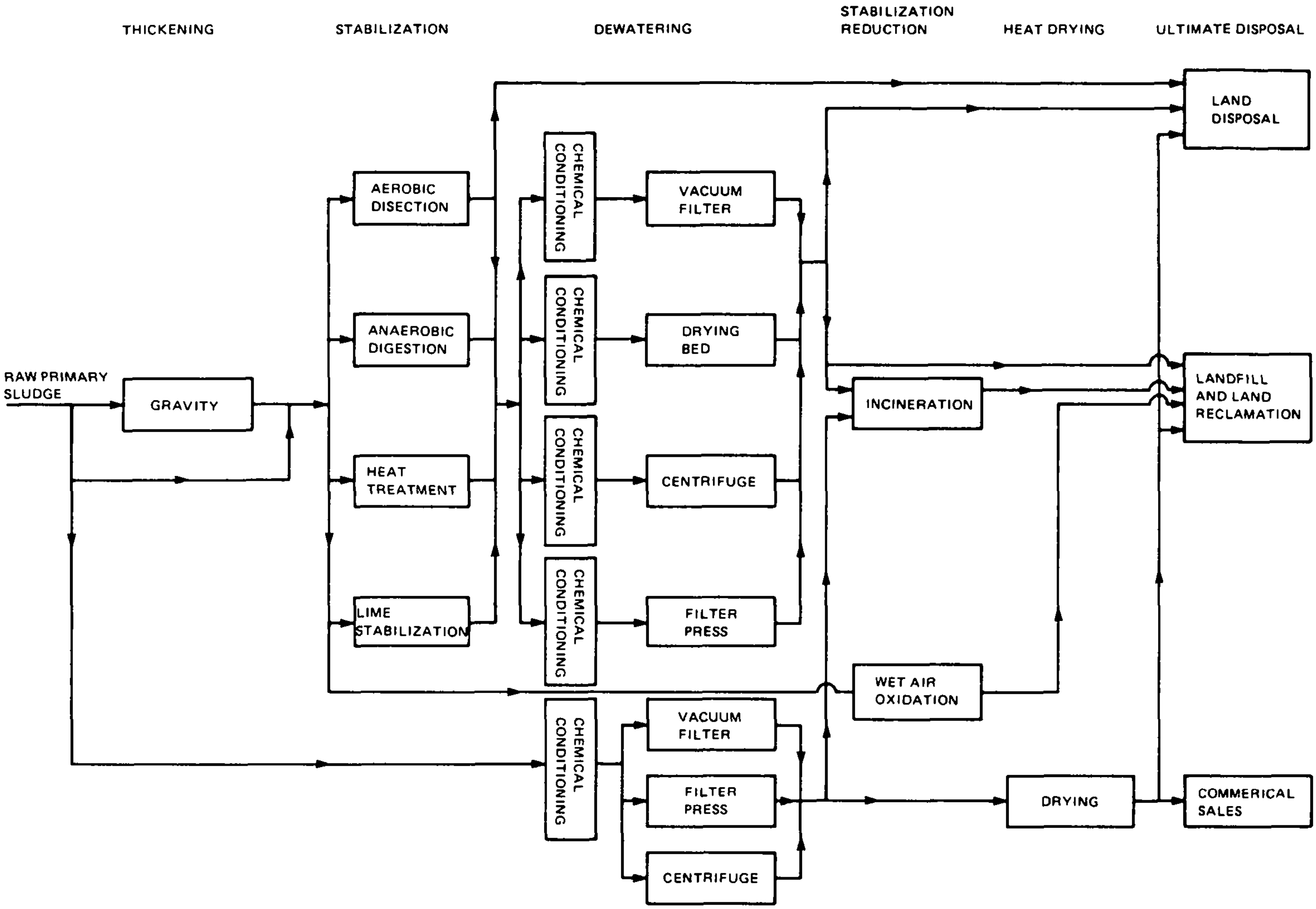


Figure 5–1.—Alternative primary sludge disposal process trains.

The lower figures in the range of expected results may be indicative of settling units processing lighter, more flocculent sludges or of units operating above their design capacity. The higher values may be indicative of the results from units processing easily settled solids or of units operating below their design capacity. Chemical additions may result in higher or lower concentrations depending upon the chemical and dosage utilized. The "typical" percent solids are indicative of the results obtained from settling tanks operating at design capacity and treating normal "domestic wastewater."

Treatment and disposal of sludges represent two of the major problems associated with wastewater treatment. Thickening of the sludge represents but one step of a total disposal scheme which may include thickening, stabilization, dewatering, stabilization reduction, or heat drying prior to ultimate disposal. Figures 5–1 and 5–2 show various primary and secondary sludge disposal alternatives and how sludge thickening may fit into the total treatment and disposal scheme.

In general, the required degree of thickening is directly related to the sludge processing method(s) downstream of the thickener (see figures 5–1 and 5–2). The stabilization stage, in particular, will normally be more successful if the solids concentration is kept within the range that optimizes the rates of biological and chemical stabilization. Likewise, ultimate disposal of liquid sludge by land application will generally be less costly when the solids concentration is maximized but kept within the range dictated by pumping equipment. Suggested optimum percent dry solids operating ranges for various sludge handling processes following thickening are shown in table 5–3.

## THICKENING PROCESSES

### Gravity Thickening

Gravity thickening of sludges, probably the most common unit process in use, is relatively simple in principle

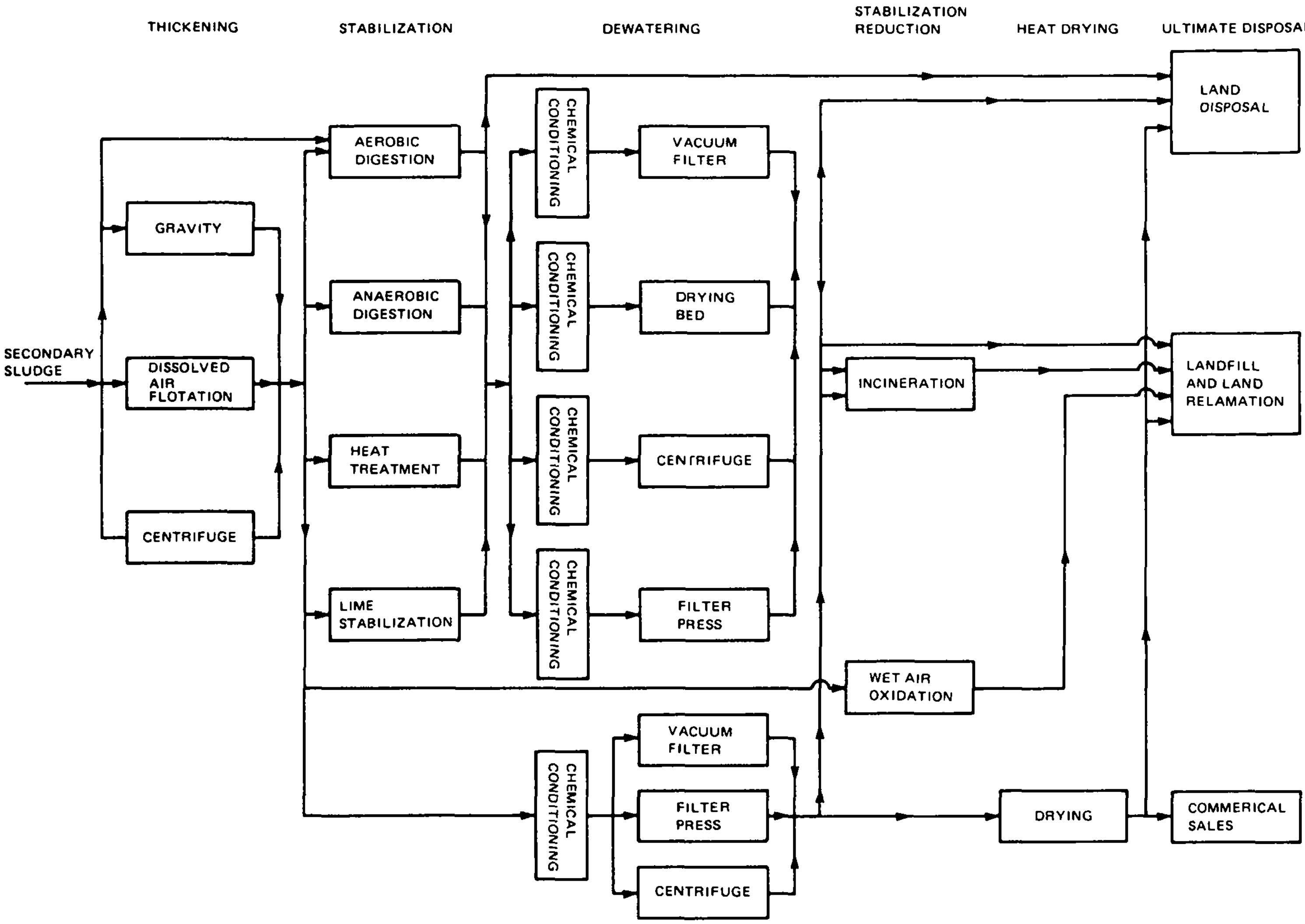


Figure 5–2.—Alternative secondary sludge disposal process trains.

Table 5–3.—Post thickening process operating ranges

| Process type | Operating ranges optimum sludge solids, percent |
|---|---|
| Stabilization | |
| Aerobic digestion | 2–4 |
| Anaerobic digestion | 4–6 |
| High pressure wet oxidation | 4–6 |
| Low pressure heat treatment | 4–6 |
| Lime treatment | 6–8 |
| Other | |
| Land application | 6–8 |

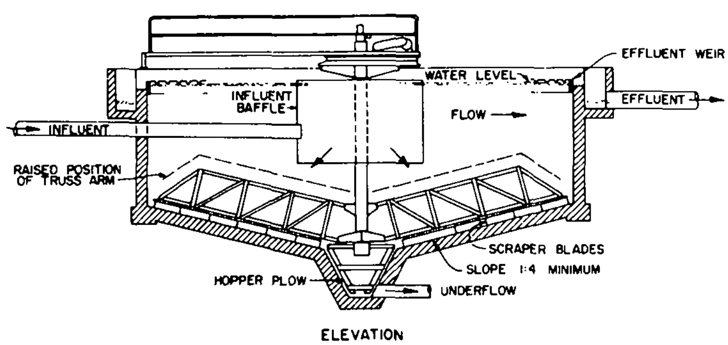


Figure 5–3.—Gravity thickener.

and operation and low in cost. Gravity thickening is basically a sedimentary process carried out in a unit which resembles a wastewater settling basin. A typical unit is shown in figure 5–3. Solids settle to the thickener bottom, are then raked to a sludge hopper, and are periodically removed and discharged to the next process. Water separated from the sludge (supernatant) rises as the sludge settles. This supernatant or overflow containing some solids and probably a high biochemical

Table 5–4.—Existing gravity thickener performance data

| Location | Feed | Sludge solids, percent | | Solids loading (lbs/ft²/day) |
|---|---|---|---|---|
| | | Unthickened | Thickened | |
| Rumford Mexico, Me. | WAS | 1.2 | 2.7 | 5 |
| Kokomo, Ind. | Heat treat[a] | 4–6 | 14–18 | 18 |
| York, Nebr. | Combined[b] | | 6–7 | |
| Salem, Ohio | PRI | | 6 | 6 |
| Middletown, Ohio | WAS | 0.9 | 3.8 | 1.5 |

[a]Contains heat treated primary and waste activated (equal portions).
[b]Contains primary, intermediate (trickling filter), and final (biodisc), proportions unknown.

oxygen demand should be returned to the plant for further treatment. Several existing gravity thickener installations were recently contacted. Data, indicative of equipment performance at that time, are presented in table 5–4.

Gravity thickeners are normally circular in shape and have a side water depth of about 10 feet (3.0 m). The tank diameter is a function of the required surface area. The required surface area is determined by applying either pilot tested or average recommended solids loading rates to the total solids that the unit will receive each day. Tank side water depth is influenced by the desired retention time and equipment availability. Sludge solids concentrations obtainable by gravity thickening depend upon the sludge type, thickener overflow rate, and solids retention time. Average recommended solids loading rates and the possible performance for some sludges are presented in table 5–5.

The values are average ranges only and may or may not be indicative of the possible results for the particular sludge in question. A case in point is a community

Table 5–5.—Gravity thickener loading rates and performance

| Sludge type | Sludge solids, percent | | Solids loading (lbs/ft²/day) |
|---|---|---|---|
| | Unthickened | Thickened | |
| Primary (PRI) | 2–7 | 5–10 | 20–30 |
| Waste activated (WAS) | 0.5–1.5 | 2–3 | 4–8 |
| Extended aeration (EA) | 1–3 | 1.5–4 | 4–8 |
| Trickling filter (TF) | 1–4 | 3–6 | 8–10 |
| Biodisc (RBD) | 1–3.5 | 2–5 | 7–10 |
| Combinations: | | | |
| PRI + WAS | 2.5–4 | 4–7 | 8–16 |
| PRI + TF | 2–6 | 5–9 | 12–20 |
| PRI + RBD | 2–6 | 5–8 | 10–18 |
| WAS + TF | 0.5–2.5 | 2–4 | 4–8 |

which gravity thickens a 0.9 percent dry solids waste activated sludge to 3.8 percent with solids capture of over 90 percent. The solids loading is 2 to 4 lbs/ft² (.91 to 1.81 kg) per day and the hydraulic loading ranges from 50 to 100 gal/ft²/day (2.0 to 4.1 m³/m²/d). This plant treats a high percentage of paper mill waste which results in significant concentrations of inorganic solids escaping the primary tanks. These solids, when combined with the biological sludge, form a floc that has much better settling characteristics than most waste activated sludges. This results in a better than average thickened product.

Although the solids loading usually governs gravity thickener design, the hydraulic loading should also be checked. Hydraulic loadings in the range of 600 to 800 gal/ft²/day (24.4 to 32.6 m³/m²/d) have been reported as optimum.[1] Also, loadings below 400 gal/ft²/day (16.3 m³/m²/d) have been reported as possibly resulting in odor problems; recycling of secondary effluent to maintain the higher rates has been recommended.[1] Much lower rates, as low as 100 to 200 gal/ft²/day (4.1 to 8.1 m³/m²/d), will often be more applicable. Recycling of secondary effluent to control odor will dilute the influent solids. The overall solids thickening performance of the unit may not deteriorate, however, since dilution will elutriate fine solids and reduce the interference between the settling particles. Polyelectrolyte addition may have the effect of improving solids capture and thus reducing solids overflow in the supernatant, but may have little effect on improving the solids concentration in the underflow. To achieve maximum sludge concentration, a sludge retention time of one-half to 2 days is normally required.

## Dissolved Air Flotation

Dissolved air flotation is presently the most widely used method of thickening waste activated sludge. The system uses air buoyancy to literally float solids to the surface of a tank to be collected. The main advantage of this method over gravity thickening is that very light particles, such as waste-activated sludge solids, can be removed more completely in less time. A typical dissolved air flotation system is shown in figures 5–4 and 5–5. The units physically range from small steel package units to custom designed large units with concrete tanks. Recycle flow may consist of either underflow from the unit or recycled plant effluent. It is returned at rates of up to five times the feed sludge rate, combined with air, and then pressurized to approximately 60–70 lbs/in.$^2$ (4.2–4.9 kg/cm$^2$). Since the solubility of air in water increases with increasing pressure, large quantities of air go into solution. Later, this recycle flow is allowed to depressurize as it is mixed with the influent sludge. Depressurization releases the excess air out of the recycle liquid in the form of tiny air bubbles (80 microns). These air bubbles attach themselves to the sludge solids and float them to the surface. Thickened sludge is scraped off the liquid surface by a skimmer mechanism consisting of a series of paddles. Liquid that is not contained in the thickened sludge or recycled is discharged from the system as subnatant. Subnatant may contain high solids and biochemical oxygen demand, and thus should be returned to the plant for further treatment.

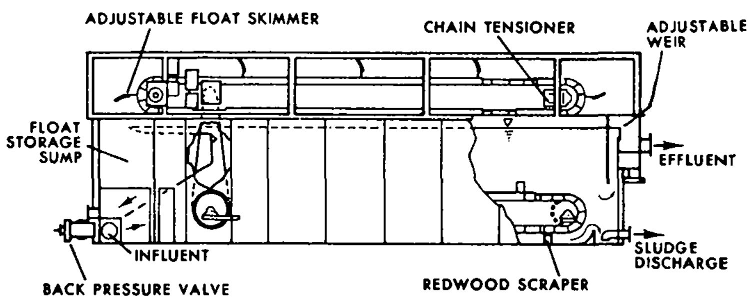


Figure 5–4.—Dissolved air flotation unit.

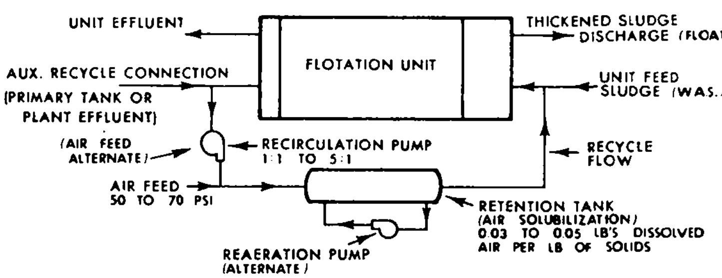


Figure 5–5.—Dissolved air flotation system.

Data from existing operating full-scale dissolved air flotation units have been presented in other publications.[2,3] Some of the same installations were recently contacted. Updated performance data for these and other dissolved air flotation units are presented in table 5–6.

The effluent sludge (float) percent solids will depend on many variables including the type and quality of the feed sludge, recycle ratio, detention time, air to solids ratio, system pressure, the solids and hydraulic loading rates, and the amount of chemical aids used. Some general statements that have been made regarding dissolved air flotation thickening of the "average" waste-activated sludge are as follows:[2]

1. Increased air pressure or flow will yield higher float solids and lower effluent suspended solids concentration.
2. Polymer usage will yield higher float solids concentration and improve the subnatant quality.
3. Detention time in the flotation zone is not critical.

Since there are so many variables and each sludge will react somewhat differently to the dissolved air flotation thickening process, these "general rules of thumb"

Table 5–6.—Recent data for some plant scale DAF units

| Location | Feed | Influent SS (mg/l) | Subnatant SS (mg/l) | Float percent solids | Polymer used lbs/ton dry solids |
|---|---|---|---|---|---|
| Indianapolis, Ind. | WAS[a] | 10,000 | 100–1,000 | 3.5–4.2 | 30 |
| Warren, Mich. | WAS[b] | 11,000 | 200 | 5 | 40 |
| Frankenmuth, Mich. | WAS | 8,000 | 90 | 3.5–5.5 | [b]0–26 |
| Columbus, Ohio | WAS[c] | 6,000 | 800 | 3.2 | 0 |
| | WAS[d] | 8,000 | | 3 | |
| Nashville, Tenn. | PS,WAS[e] | 35,000;5,000 | 150 | 6 | 0 |
| Xenia, Ohio | WAS | 4,000 | 100 | 2.5–3.0 | 30 |

[a]Contains some primary sludge—proportions unknown.
[b]Major flow to plant is brewery waste. Polymer sometimes used to keep sludge from adhering to skimmers. Sometimes thicken anaerobically digested sludge—similar results with no polymers required (influent SS 10,000 mg/l).
[c]Jackson Pike facility.
[d]Southerly facility—units are being used as gravity settlers since they get better results this way.
[e]Primary and waste activated are handled by separate units—combined product is 6 percent solids.

Table 5–7.—Dissolved air flotation design parameters and expected results

| Sludge type | Feed solids, percent | Solids loading ($lb/ft^2/hr$) | Air to solids ratio | Recycle ratio, percent | Float solids, percent | | Solids capture, percent | |
|---|---|---|---|---|---|---|---|---|
| | | | | | With polymer | Without polymer | With polymer | Without polymer |
| Waste activated | 0.5–1.5 | 2–3 | 0.03–0.05 | 100–500 | 5–6 | 4–5 | 95–100 | 85–95 |
| Primary and waste activated | 3–4 | 2–4 | [a]— | [a]— | [a]— | 5–8 | [a]— | 85–95 |

[a]Limited experience prohibits listing typical numbers.

may not apply in all cases. Additionally, when the guidelines are valid, it is generally only within certain ranges of the variable parameters. The ranges are typically 40–70 $lbs/in.^2$ ($2.8–4.9 kg/cm^2$) for air pressure and 0–40 lbs (0–18.1 kg) for polymer dosage. Likewise, the detention time may not be critical once a minimum value of 1.5–3 hours has been attained.

System design is based primarily on a solids loading rate and the desired air to solids ratio. Additionally, maximum hydraulic loading rates are usually checked to avoid exceeding manufacturers' recommendations. If any flow other than the dissolved air flotation thickener underflow is used for recycle, it must be included in the unit's total hydraulic loading calculation.

Pilot studies are recommended to determine the applicability of the dissolved air flotation process to the sludge and to optimize some of the variables. When pilot studies are undertaken, the full-scale design is based on the study findings. Since data and sludge samples are lacking at new wastewater treatment plants, thickener design must be based on sound engineering judgment backed up with past experience. Commonly used design parameters and expected unit performance are presented in table 5–7. It must be emphasized that these are general guidelines only.

## Centrifugation

Centrifugal thickening of sludge is a process which uses the force developed by fast rotation of a cylindrical drum or bowl to separate the sludge solids and liquid. In the basic process, when a sludge slurry is introduced to the centrifuge, it is forced against the bowl's interior walls, forming a thin slurry layer or "pool." Density differences cause the sludge solids and the liquid to separate into two distinct layers. The sludge solids "cake" and the liquid "centrate" are then drawn from the unit separately and discharged. The three types of centrifuges—basket, disc-nozzle, and solid bowl—all operate on the basic principles described above. They are differentiated by the method of sludge feed, applied centrifugal force, method of solids and liquid discharge, and to some extent performance.

The basket centrifuge, as shown in figure 5–6, is a

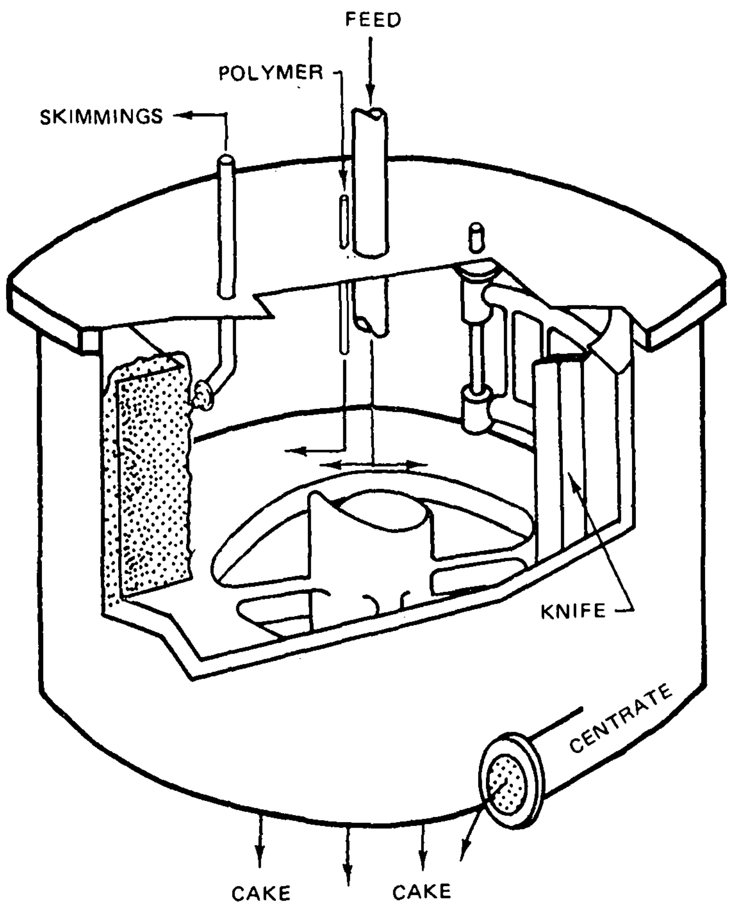


Figure 5–6.—Schematic diagram of a basket centrifuge.

batch type thickening unit. As slurry is fed to the unit, the sludge solids form a cake on the bowl walls, while the centrate is discharged over a weir or baffle. Slurry feed is continued until the centrate solids reach the maximum tolerable limit. At this point, the unit stops and a knife wipes the sludge cake off the walls. The sludge is then discharged from the system through the unit's open bottom. Of the three centrifuge types, the basket unit has the capability of producing the driest sludge cake since there is a minimum of disturbance to the depositing solids. Its use, however, is generally restricted to smaller plants because of its intermittent operation and resultant lower capacity.

The disc-nozzle centrifuge, as shown on figure 5–7, is a continuously operating unit. It is composed of a series of conical plates which are stacked together to form a series of narrow channels. Sludge slurry enters the unit and is dispersed to these channels. The centrate tends to rise and is discharged from the top of the cones, while the sludge cake is discharged downward and through small nozzles in the bowl wall at the cone bottoms. High sludge throughput and good solids capture are possible with these units. Their solids concentrating capability is limited, however, by the small diameter (0.05–0.10 in.) (0.13–0.25 cm) orifices through which the sludge cake must discharge. Additionally, depending upon the sludge type and previous treatment, degritting and screening prior to the disc centrifugation may be mandatory to avoid plugging these sludge discharge orifices and to reduce wear on the machine.

Like the disc centrifuge, the continuous solid bowl centrifuge is a continuously operating unit. It consists of a horizontal cylindrical bowl containing a screw type conveyor. At one end, the bowl necks down to a conical section that acts as a beach plate for the screw conveyor. In operation, sludge solids are forced to the bowl surface and are moved toward the beach plate by the conveyor where they are discharged from the unit. The sludge pool level is controlled by adjustable skimmers or weir plates. These also function as discharge points for the centrate. A typical countercurrent solid bowl centrifuge is shown in figure 5–8. Sludge slurry enters the unit just before the conical section and distributes itself along the bowl surface. Sludge solids are discharged at the cone end while centrate is discharged at the opposite end. Sludge solids do not travel the full length of the bowl. A second variation of the solid bowl centrifuge is the concurrent model. In this unit, sludge slurry is introduced at the far end of the bowl. Turbulence and interference present at the slurry inlet point in the countercurrent machine are reduced with this variation. Also, the slurry must travel the full length of the bowl before discharge. This may result in a drier sludge cake.

Centrifuge performance is measured by the percent solids of the sludge cake and the centrate quality or total solids captured. Several existing centrifuge installations were recently contacted. Data, indicative of equipment performance at that time, are presented in table 5–8. The performance of a particular centrifuge unit will vary with the inlet sludge type and solids characteristics, the sludge feed rate, and the degree of chemical addition. Centrifuge performance on a particular sludge will also vary with bowl design, bowl speed, pool volume, and conveyor (if present) design. In practice, bowl and conveyor design are set by the manufacturers. Pool depth is variable on solid bowl units. Increasing the pool depth will normally result in a wetter sludge cake but better solids recovery. Bowl speed is normally variable on most centrifuge models. Difficulty involved in changing speeds varies with the manufacturers. An increase in bowl speed normally results in a drier sludge cake and better solids recovery. Conveyor differential speed is normally variable on continuous solid bowl centrifuges. Increasing the differential normally results in a wetter sludge cake and poorer solids recovery. Varying these parameters will probably result in significant solids

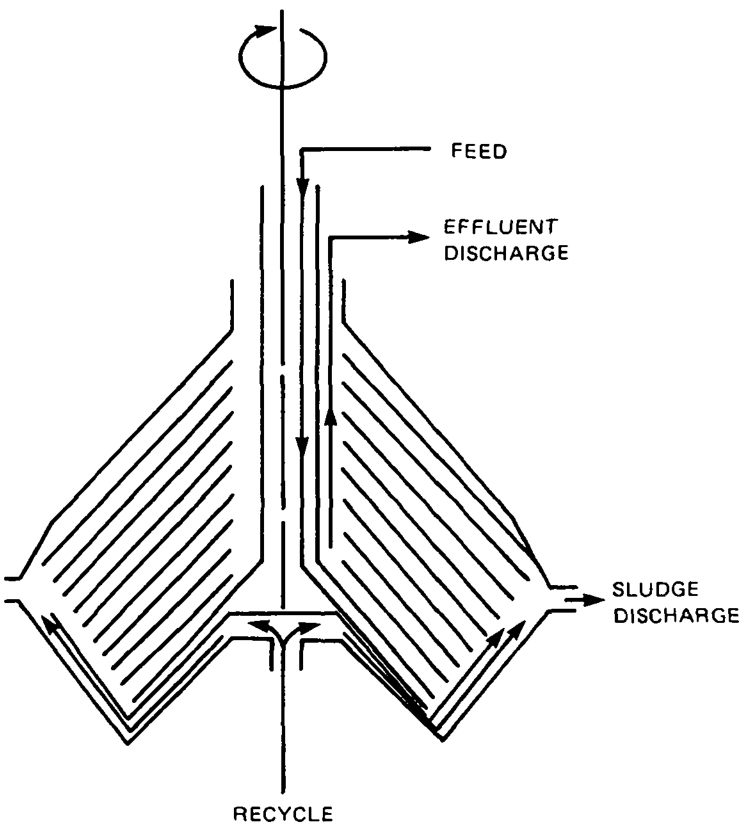


Figure 5–7.—Disc-nozzle centrifuge.

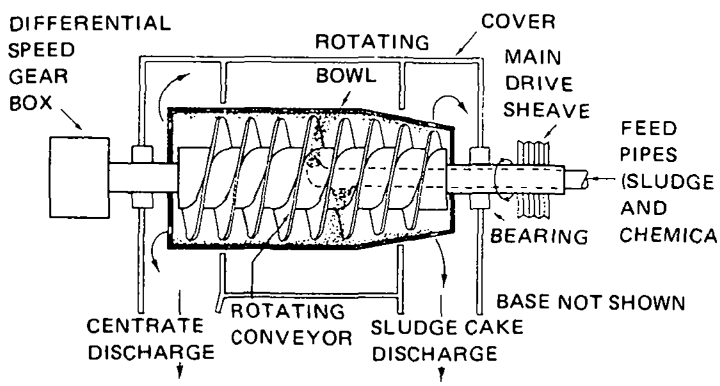


Figure 5–8.—Continuous countercurrent solid bowl conveyor discharge centrifuge.

Table 5–8.—Existing solid bowl centrifuge performance data

| Location | Feed | Sludge solids, percent | | Solids recovery, percent |
|---|---|---|---|---|
| | | Unthickened | Thickened | |
| Great Northern Paper, Millinocket, Maine | WAS[a] | 4 | 10–12 | 90 |
| Kendall Co., Griswoldville, Mass. | WAS[a] | 3 | 7 | |
| Miller Brewing Co., St. Louis, Mo. | WAS[b] | 0.75–1 | 5–7 | 80–85 |
| Dubuque, Iowa | WAS[b] | 1–1.5 | 6 | |

[a]Polymers used—quantity unknown.
[b]Polymers not used.

Table 5–9.—Centrifuge mechanical characteristics and performance data

| Parameter | Centrifuge type | | |
|---|---|---|---|
| | Basket | Disc-nozzle | Solid bowl |
| Operation method | Batch | Continuous | Continuous |
| Bowl diameter (inches) | 12–60 | 8–30 | 6–60 |
| Max centrifugal force (G) | 2,000 | 12,000 | 3,200 |
| WAS feed solids, percent | 0.5–1.5 | 0.5–1.5 | 0.5–1.5 |
| WAS cake solids, percent | 8–10 | 4–6 | 5–8 |
| Solids recovery, percent | 80–90 | 80–90 | 70–90 |

changes only within limited ranges. Each performance improvement must be compared with the additional costs required to produce it.

Centrifuges have seen more service in dewatering applications than in thickening applications. When utilized for thickening, their use is normally limited to the thinner biological or industrial sludges that cannot be thickened by less expensive methods. Data on the three centrifuge types and their possible performance on waste activated sludge are presented in table 5–9.

Polymers may be required to meet the stated performance. The required dosage depends upon the manufacturer and may range from 0–8 lb/ton (0–4.0 kg/Mg) of dry solids.

Polymer addition generally improves both the percent solids and the solids recovery. It must be emphasized that the tabular values are representative of possible results from an "average" waste activated sludge. Numerous sludge and machine variables make consultation with manufacturers mandatory and pilot tests highly recommended for each installation.

## Other Methods

Thickening of sludge is often a secondary benefit of a sludge treatment unit having an entirely different purpose. Decanting facilities are provided in aerobic and anaerobic digesters to remove excess liquids which have risen above the solids layer. In such facilities, sludge solids concentrations may increase as much as one percent over inlet feed solids concentrations.

New sludge thickening methods are being marketed each year. One such method is the sludge filter bag system. In this process, sludge is mixed with polymer and then held in suspended porous bags. The weight of the sludge forces water out the bag sides and bottom. Sludge is held from four to eight hours depending upon the desired dryness and is then released through a bottom opening. Bag life should be about 2 years. This method has not been in existence long enough to have been proven reliable.

# DESIGN EXAMPLE

## Statement of Problem

The problem is to provide sludge thickening facilities for two communities, both of which have existing conventional activated sludge wastewater treatment plants.

The smaller community has existing wastewater treatment facilities capable of treating 4.0 million gallons per day (.18 $m^3/s$). The facilities consist of screening, grit removal, primary settling, conventional activated sludge aeration, final settling, chlorination, and sludge lagooning. Present flow to the plant is 3.5 million gallons per day (.15 $m^3/s$); the 20 year projected flow is 4.0 million gallons per day (.18 $m^3/s$). The plant meets its proposed discharge permit requirements, but the city has been ordered to abandon the sludge lagoons (which are periodically flooded by the receiving stream) and in their place construct digestion facilities and devise a plan for disposal of the digested sludge. The digested sludge will be dewatered on sand drying beds or hauled as a liquid to nearby farms. Thickening facilities are required to reduce the size of the required anaerobic digester, to insure efficient digester operation, and reduce hauling costs.

The larger community has existing wastewater treatment facilities capable of treating 30 million gallons per day (1.31 $m^3/s$). Present flow to the plant is 35 million gallons per day (1.53 $m^3/s$); the 20-year projected flow is 40 Mgal/d (1.75 $m^3/s$). The existing treatment system consists of screening, grit removal, primary settling, conventional activated sludge aeration, final settling, chlorination, aerobic sludge digestion, sludge dewatering, and landfilling of dried sludge solids. The existing treatment scheme will meet proposed permit requirements. Therefore, all treatment units will be expanded to handle the 20-year flow projections. Anaerobic digestion has been determined to be more cost-effective than the aerobic sludge digestion. The aerobic digesters will be abandoned as such (will become part of expanded aeration tank facilities). Thickening facilities are required to reduce the size of the required anaerobic digesters, to insure efficient digester operation, and to improve the dewatering operation.

### *Wastewater Characteristics*

The wastewater characteristics and removal efficiencies of the various treatment units are required to determine the possible solids loading on the thickeners. This information may be acquired from plant records or sampling programs at existing facilities. When these data are not available (such as in the case of new wastewater treatment plants for new service areas), assumptions based on sound engineering judgment and previous experience are necessary. For the sake of simplicity, the wastewater characteristics and treatment unit removal efficiencies for the example plants are assumed equal. Raw wastewater characteristics for the example plants are given in table 5–10.

### *Treatment Unit Efficiencies*

Both plants in this example will meet their proposed permit requirements by utilizing the existing treatment processes. Nitrification and phosphorus removal are not required. Removal efficiencies based on percentages of the raw "domestic" wastewater characteristics are presented in table 5–11.

### *Sludge Characteristics*

The characteristics of sludge discharged to the thickening facilities may vary considerably depending upon the type and amount of industrial wastes treated, the sludge origin (which particular treatment unit), the degree of chemical addition for improved settling or phosphorus removal, and the sludge age. Ideally, samples of the sludge will be available for analysis. In lieu of this, the ranges and typical concentrations shown in table 5–2 may be utilized.

Table 5–10.—Raw wastewater characteristics

| Parameter | Concentration (mg/l) |
|---|---|
| $BOD_5$ | 200 |
| Suspended solids | 240 |
| Organic nitrogen | 15 |
| Ammonia nitrogen | 25 |
| Phosphorus | 10 |
| Grease | 100 |

Table 5–11.—Treatment unit efficiencies

| Unit | Parameter | Removal efficiency, percent |
|---|---|---|
| Primary settling | $BOD_5$ | 30 |
| | SS | 65 |
| Aeration and final settling | $BOD_5$ | 60 |
| | SS | 25 |

Existing plant operating data at the example plants have shown that the unthickened primary sludge contains four percent dry solids; the waste activated sludge, one percent dry solids. Field experiments at both plants were conducted by returning the waste activated sludge to the primaries. This did not seriously alter their operational characteristics and an unthickened primary sludge containing 3 percent dry solids resulted. Additionally, data at these plants have shown that for every pound of 5-day biochemical oxygen demand removed in aeration, 0.5 pound of volatile suspended solids is produced.

### *Sludge Handling Following Thickening*

The required degree of thickening is directly related to the sludge processing method(s) following thickening. Suggested optimum percent dry solids operating ranges for some sludge handling processes following thickening were presented in table 5–3. In the examples, anaerobic digestion is to follow the thickening step. Hence, sludge delivered to the digester should have a solids concentration between 4 and 6 percent.

For any sludge thickening problem, there will be several alternative solutions which will result in a sludge product in the desired solids range. However, since each solution will probably not result in the same "guaranteed average" percent dry solids, the design of the sludge processing facilities following thickening will also be affected. Consequently, these facilities will also have to be included in the cost analysis.

## Process Alternatives

### *Gravity Thickening*

In the example, a primary (4 percent) and waste activated sludge (1 percent), or combined sludge (3 percent) is obtained, and a sludge concentration for the anaerobic digester of 4 to 6 percent is needed. Table 5–5 and past experience indicate that gravity thickening of "normal" waste activated sludge alone will not yield the required 4 percent solids. Gravity thickening may yield reasonable results for the combined sludges. Additionally, gravity thickening primary sludge alone and waste activated alone, and later mixing the two, is a possibility. At this point in an actual problem at an existing treatment plant, bench or pilot studies would be performed to determine the applicability of gravity thickening to the sludge and to determine design parameters.

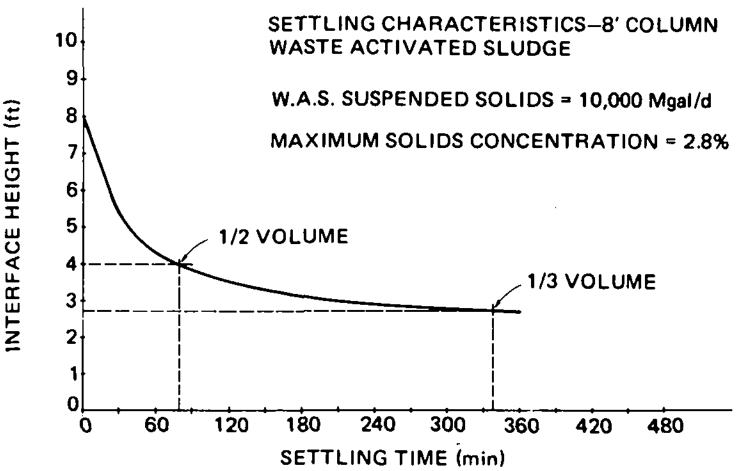


Figure 5–9.—Settling characteristics—8 foot column waste-activated sludge.

Examples of results of typical 8-foot column bench scale tests are shown on figures 5–9 and 5–10. Both the undiluted and elutriated activated sludges reached their maximum solids concentrations of 2.8 percent and 2.3 percent, respectively, in less than 3 hours. A similar test would be made on primary only and combinations of primary and waste activated sludge.

For the example plants, assume the results of the tests showed that gravity thickening the sludges will result in the following: primary sludge, nine percent; waste activated sludge, 2.8 percent; combined primary and waste activated sludge, 5 percent.

## Dissolved Air Flotation

Reviewing the example problem, there is primary (4 percent) and waste activated sludge (1 percent) or combined sludge (3 percent), and a sludge concentration for

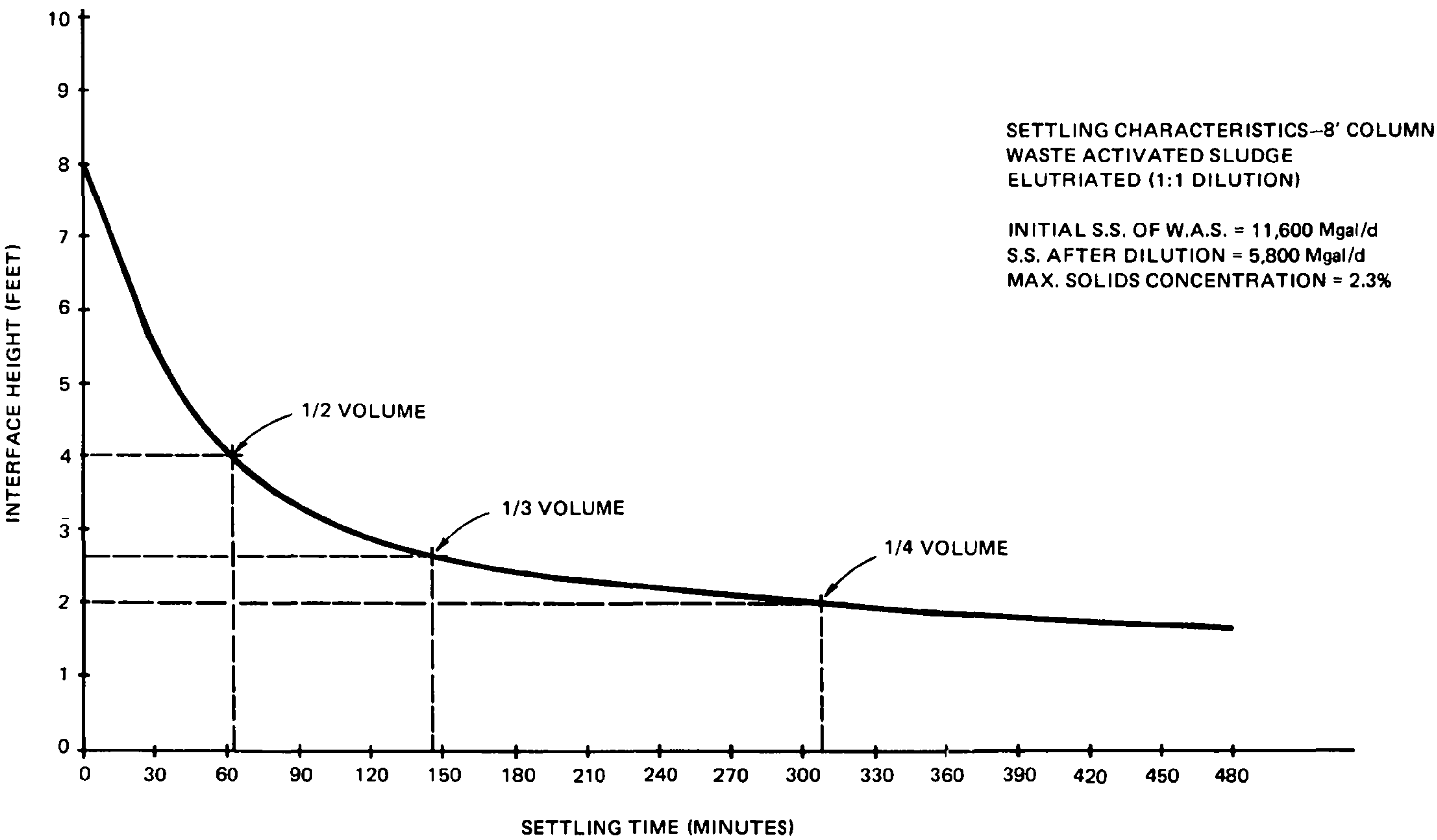


Figure 5–10.—Settling characteristics—8 foot column elutriated waste-activated sludge.

the anaerobic digesters of 4 to 6 percent is needed. If primary sludge is to be thickened alone, gravity thickening is generally utilized since the costs would be much less than for dissolved air flotation. Likewise, in the case of the primary-waste activated combined sludge, gravity thickening will yield similar results at less cost. This leaves thickening the waste activated sludge alone by dissolved air flotation thickening as a possible option. Dissolved air flotation thickening of the waste activated sludge, coupled with either unthickened or gravity thickened primary sludge, represents a viable alternative and will be considered. At existing plants, pilot tests should be performed to aid in process selection and equipment design.

Assume a pilot study was completed using dissolved air flotation thickening on the waste activated sludge. The variables studied included recycle ratio, air to solids ratio, solids loading rate, and amount of polymer used. The system pressure was kept constant. The results, shown graphically in figures 5–11, 5–12, 5–13, and 5–14, were as follows:

1. Increasing the recycle rate generally yielded higher percent float solids but also higher effluent suspended solids. A compromise rate was selected for use in later tests.
2. A concentrated sludge of 4 percent solids could be consistently achieved with a unit loading of 2 $lb/ft^2/hr$ (9.8 $kg/m^2/hr$) and an air to solids ratio of 0.04. Increasing the solids loading reduced the float concentration and increased the effluent suspended solids concentration with and without polymer usage.
3. At the recommended loading, an effluent suspended solids concentration of 50 milligrams per liter without the use of polymers and 20 milligrams per liter with polymer addition was consistently achieved. Polymer usage, however, resulted in no clearly identifiable improvement in the float solids concentration.
4. Very rapid deterioration in the effluent quality occurred when the air to solids ratio fell below 0.020. Increasing the air to solids rates from 0.040 to 0.250 resulted in only slight reduction in effluent suspended solids.

As seen from the results, the waste activated sludge differed somewhat from the experience of others[2] and an

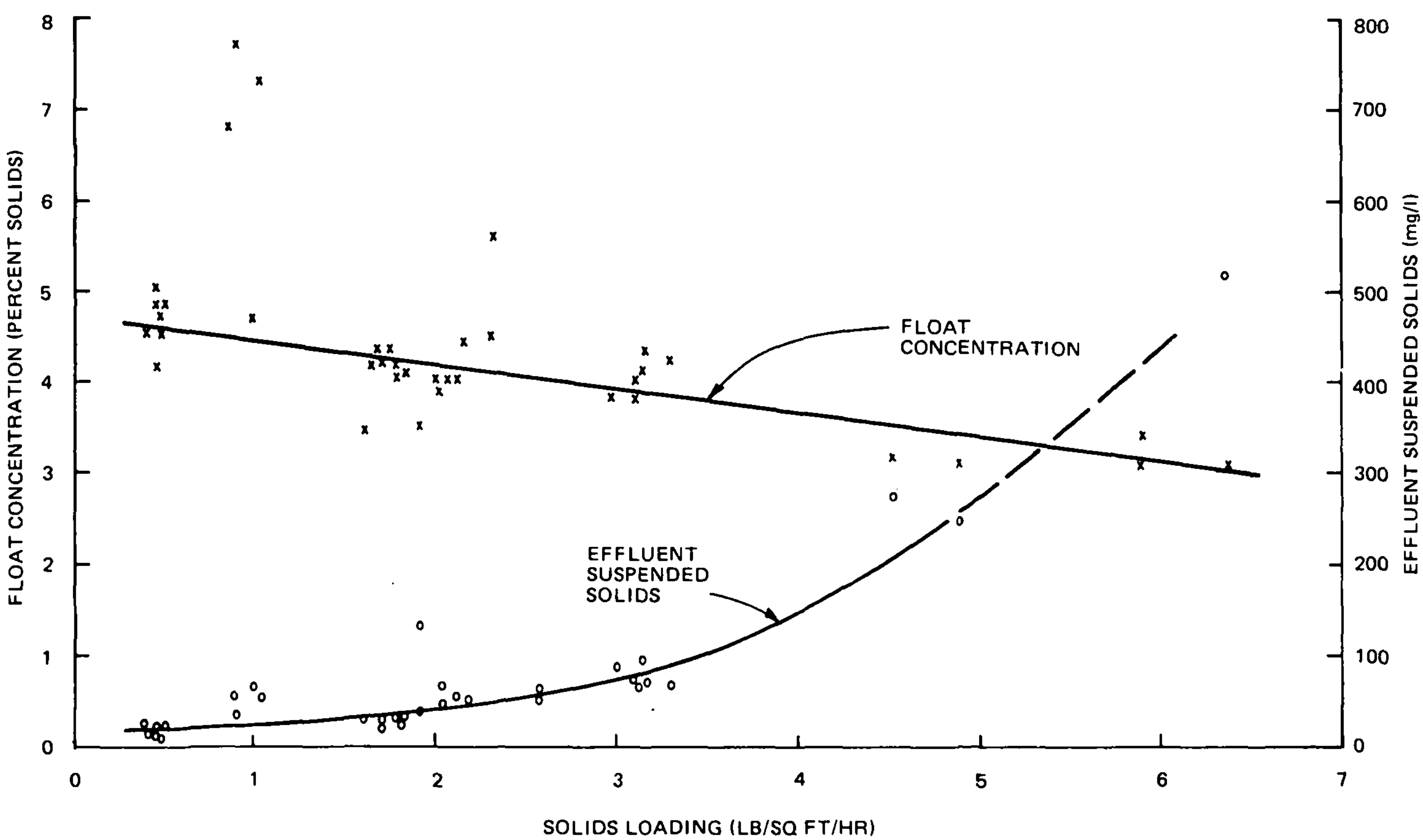


Figure 5–11.—Float concentration and effluent suspended solids versus solids loading—without polymers.

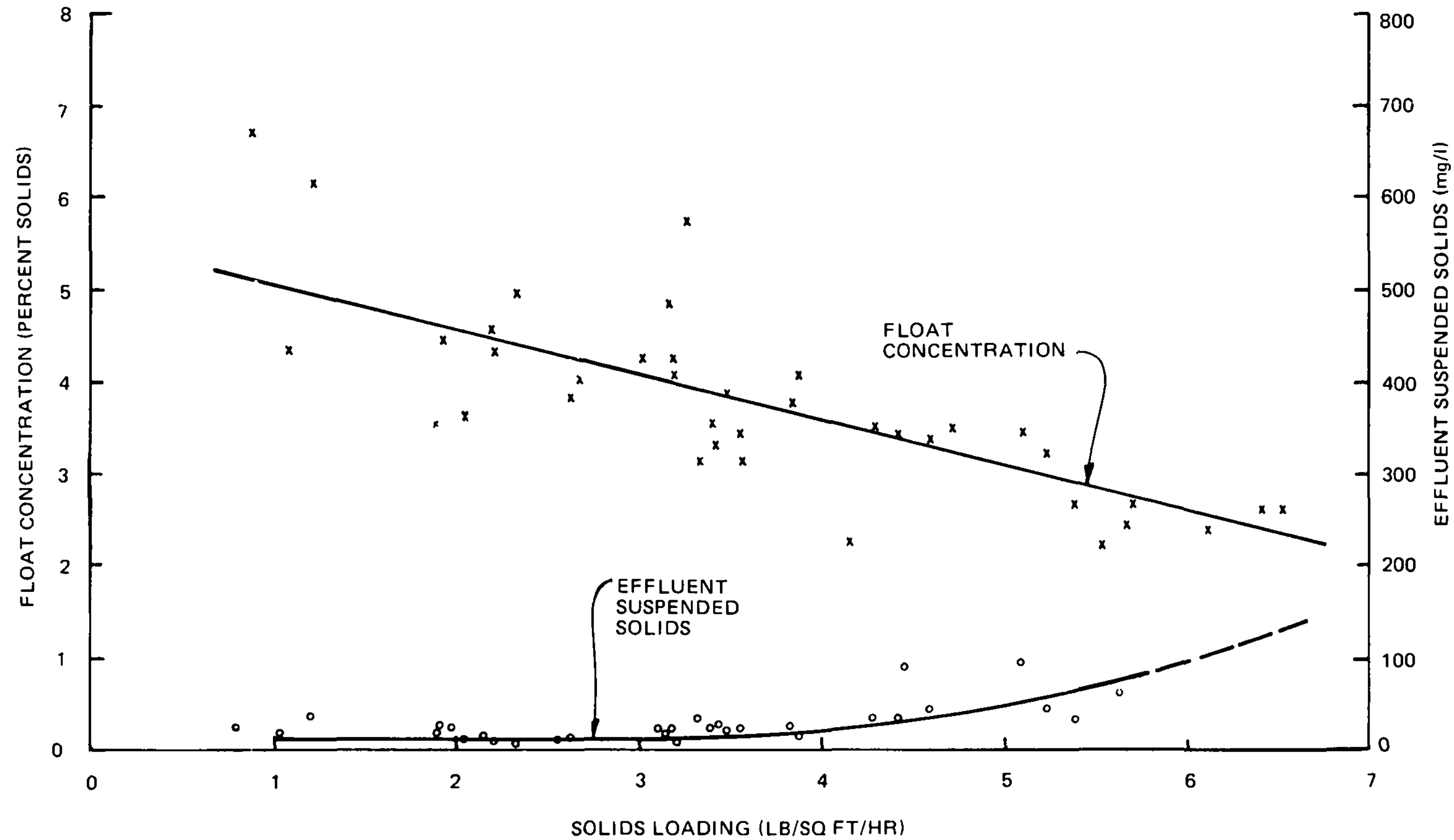


Figure 5–12.—Float concentration and effluent suspended solids versus solids loading—with polymers.

"average" waste activated sludge. A 4 percent float was obtained with or without polymers.

For the example plants, it will be assumed that dissolved air flotation thickening is applicable to the waste activated sludge and that a thickened sludge of 4 percent solids will be produced at solids loadings of 2 $lb/ft^2/hr$ (9.8 $kg/m^2/hr$) and an air to solids ratio of 0.04.

## Centrifugation

The problem at the example wastewater plants is to produce a 4 to 6 percent dry solids sludge for anaerobic digestion from primary sludge (4 percent) and waste activated sludge (1 percent), or combined sludge (3 percent). Past experience indicates that thickening the primary or the combined sludge by centrifugation would be a more costly alternative than gravity thickening. These alternatives are eliminated from further consideration. Centrifugal thickening of the waste activated sludge, however, combined with either unthickened or gravity thickened primary sludge does represent a viable alternative and will be considered. As in the case of gravity and dissolved air flotation thickening, sludge treatability and variable optimization make pilot studies highly desirable when possible.

For the example, assume a pilot study using a solid bowl centrifuge was performed as part of the sludge thickening study on the waste activated sludge. Some typical data from this pilot test are shown in table 5–12. In the pilot study, the feed rate of the sludge, bowl speed, and pond setting were varied to determine the optimum combination to yield a 5 percent sludge. Minor pond setting changes had little effect on the unit's performance. Operation at 3,200 G produced a sludge much thicker (12 percent) than needed, while operation at 1,150 G produced a wet sludge and poor solids removal efficiency. A force of 2,100 G was selected as an optimum. At the selected bowl speed, solids recovery and percent solids of the cake were analyzed for different sludge feed rates. The data indicated that although the centrifuge could thicken the sludge to the required 5 percent, the percent solids could drop from 5 percent down to 2 percent or increase up to 15 percent, with only minor feed rate changes. Consistently obtaining the required 5 percent solids concentration was difficult. Based on the pilot test data, solid bowl centrifuge thickening of the waste activated sludge was not consistent.

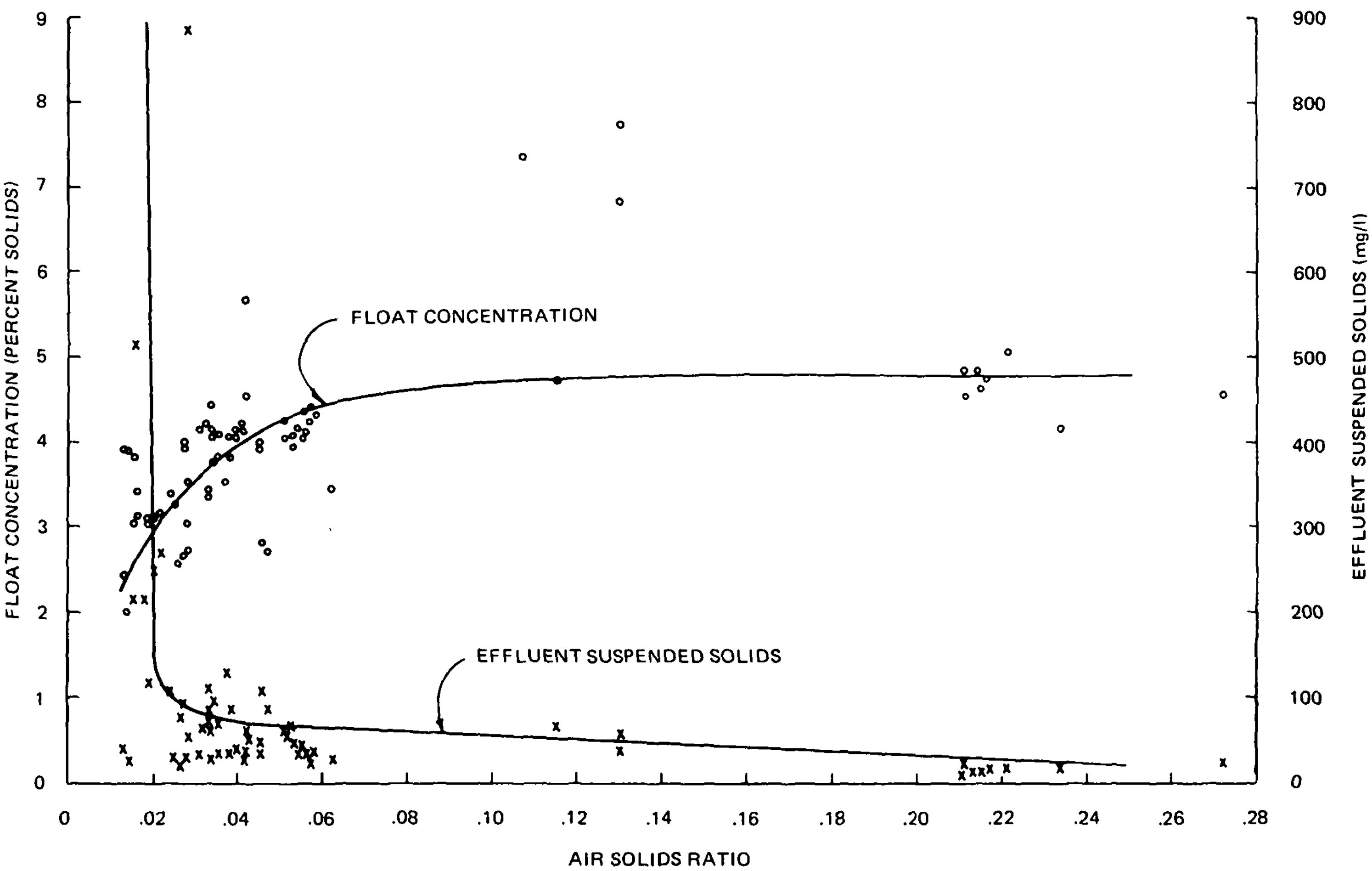


Figure 5–13.—Float concentration and effluent suspended solids versus air-solids ratio—without polymers.

For the example plants, however, it will be assumed that centrifugation is applicable to the waste activated sludge. Also, based on available equipment reliability, plant operator preference, desired performance, minimum supportive equipment requirements, and past experience, the solid bowl continuous centrifuge is selected over the basket and disc centrifuge for the examples. Available data from equipment manufacturers and data in table 5–9 indicate that a product sludge of 6 percent solids may be reasonably expected.

### *Other Methods*

Decanting may result in some thickening in the digesters. It is not, however, a reliable, consistent method and does not normally result in appreciable thickening. Thus, it will not be considered as one of the process alternatives for the example plants.

New methods, such as the sludge filter bag system, have not been in existence long enough to have been proven reliable. Thus, they will not be considered as thickening process alternatives for the example plants.

## Alternative Evaluation

### *Preliminary Screening*

The preliminary screening of sludge thickening alternatives for the example plants was performed in the previous section. The remaining alternatives at this point are presented in table 5–13.

The general approach to use, at this point, is to first determine if any of the remaining alternatives can be eliminated without performing a detailed cost-effectiveness analysis. A detailed cost-effectiveness analysis examining capital and operation and maintenance costs would then be performed on the remaining alternatives. Capital costs to be considered may normally include thickener and supportive equipment costs, land costs, building or protective structure costs, and, in certain cases, post thickening treatment unit costs. Other costs to be considered include power costs, chemical costs, manpower costs, and maintenance costs. The cost-effectiveness analysis will show which alternative has the lowest annual equivalent cost.

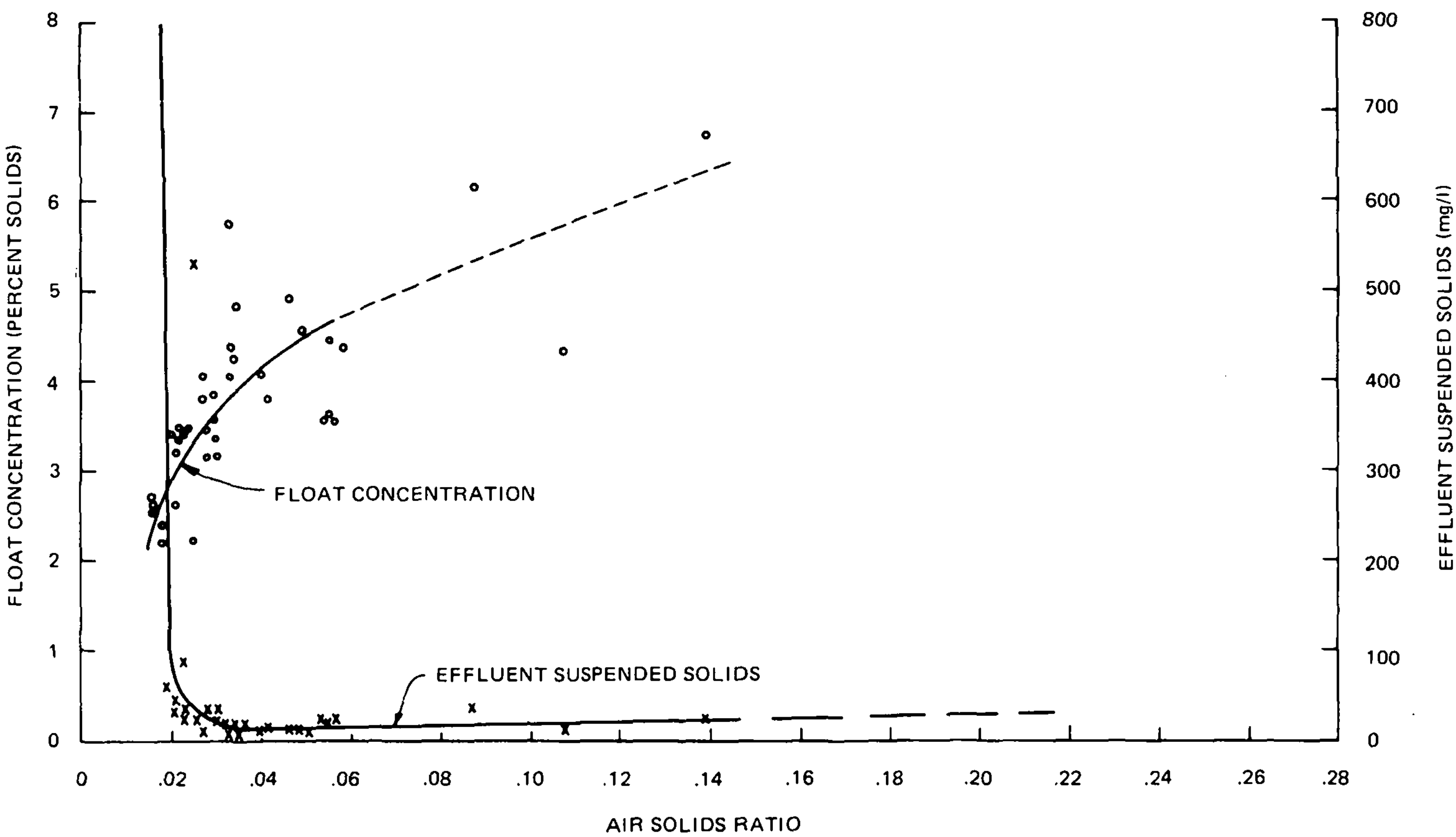


Figure 5–14.—Float concentration and effluent suspended solids versus air-solids ratio—with polymers.

## *Secondary Screening Analysis*

Since alternative numbers 1 and 2 both utilize gravity thickening only, elimination of one of them should be relatively simple. Wastewater characteristics and settling tank performance data presented previously will be used in determining loadings on the required thickeners. For the examples, differences in density of the sludges are assumed insignificant and the density is taken as equal to water. Thickener designs will be based on loading rates proposed in tables 5–5, 5–7, and 5–9. Designs will be conservative to assure the desired performance. A total of two thickeners will be provided with each alternative to assure that some thickening will be obtained if one unit fails. Calculations required for the 4.0 million gallons per day wastewater plant gravity thickener designs follow:

### Alternative No. 1

Definition—Gravity thicken primary sludge; gravity thicken waste activated sludge.

Primary sludge

Quantity: 4 × 240 × 8.34 × 0.65 = 5,204 lbs/day (2360 kg/day)
Volume: 5,204/(0.04 × 8.34) = 15,600 gals/day (59,050 l/day)
Required thickener: 5,204/20 lb/ft$^2$/day = 260 ft$^2$ (24.2 m$^2$) or an 18.2 ft (5.55 m) dia. unit
Recommended thickener: one 20 ft (6.10 m) dia., 10 ft (3.05 m) deep unit
Thickened product: 5,204/(0.09 × 8.34) = 6,933 gals/day (26,240 l/day)
Thickener cost: $64,000

Waste activated sludge

Nonbiological: 4 × 240 × 8.34 × 0.25 = 2,002 lbs/day (908 kg/day)
Biological: 4 × 200 × 0.60 × 8.34 × 0.5 = 2,002 lbs/day (908 kg/day)
Total quantity: 4,004 lbs/day (1816 kg/day)
Volume: 4,004/(0.01 × 8.34) = 48,010 gals/day (181,740 l/day)
Required thickener: 4,004/4 lbs/ft$^2$/day = 1,001 ft$^2$ (93.0 m$^2$) or a 35.7 (10.88 m) dia. unit
Recommended thickener: one 35 ft (10.67 m) dia., 10 ft (3.05 m) deep unit
Thickened product: 4,004/(0.028 × 8.34) = 17,146 gals/day (64,905 l/day)
Thickener cost: $98,000

Combined product

[(6,933 × 9) + 17,146 (2.8)]/(6,933 + 17,146) = 4.59
24,079 gals/day (91,950 l/day) of 4.59 percent sludge

### Alternative No. 2

Definition—Gravity thicken combined sludge.

Combined sludge

Nonbiological: 4 × 240 × 8.34 × 0.9 = 7,206 lbs/day (3269 kg/day)
Biological: 2,002 lbs/day (908 kg/day)
Total quantity: 9,208 lbs/day (4177 kg/day)
Volume: 9,208/(0.03 × 8.34) = 36,803 gals/day (139,315 l/day)

Table 5–12.—Pilot centrifuge results

| Run No. | Feed sludge Rate gpm | Feed sludge Concentration percent SS | Centrate Rate gpm | Centrate Concentration percent SS | Cake Concentration percent SS | Cake Percent solids recovered | Mechanical conditions Bowl speed, rpm | Mechanical conditions Pond setting | Mechanical conditions Bowl-conveyor differential (rpm) |
|---|---|---|---|---|---|---|---|---|---|
| 1 ..... | 13.6 | 0.799 | 12.5 | 0.027 | 9.7 | 97 | 3,250 | 8-3/4 | 4.0 |
| 2 ..... | 10.8 | .859 | 6.8 | .018 | 2.3 | 99 | 3,250 | 8-3/4 | 9.2 |
| 3 ..... | 16.8 | .817 | 15.8 | .077 | 11.5 | 91 | 3,250 | 8-3/4 | 3.1 |
| 4 ..... | 17.7 | .925 | 16.2 | .034 | 10.7 | 96 | 3,250 | 8-3/4 | 4.2 |
| 5 ..... | 18.3 | .918 | 10.0 | .020 | 2.0 | 99 | 3,250 | 8-3/4 | 7.3 |
| 6 ..... | 25.2 | .833 | 24.0 | .230 | 12.7 | 73 | 3,250 | 8-3/4 | 3.3 |
| 7 ..... | 25.7 | .845 | 24.0 | .072 | 10.9 | 92 | 3,250 | 8-3/4 | 4.8 |
| 8 ..... | 21.6 | .809 | 13.0 | .024 | 2.0 | 98 | 3,250 | 8-3/4 | 6.3 |
| 9 ..... | 23.0 | .813 | 17.2 | .027 | 3.1 | 96 | 3,250 | 8-3/4 | 2.8 |
| 10 ..... | 35.4 | .809 | 22.2 | .039 | 2.1 | 97 | 3,250 | 8-3/4 | 2.8 |
| 11 ..... | 23.6 | .782 | 22.2 | .136 | 11.5 | 86 | 3,250 | 8-5/8 | 4.1 |
| 12 ..... | 12.2 | .790 | 10.0 | .018 | 4.3 | 98 | 3,250 | 8-5/8 | 5.9 |
| 13 ..... | 10.7 | .699 | 10.0 | .015 | 10.5 | 98 | 3.250 | 8-5/8 | 4.1 |
| 14 ..... | 11.1 | .757 | 9.7 | .014 | 5.8 | 98 | 3,250 | 8-5/8 | 5.4 |
| 15 ..... | 22.2 | .757 | 20.0 | .026 | 7.5 | 97 | 3,250 | 8-5/8 | 5.4 |
| 16 ..... | 27.3 | .779 | 26.1 | .191 | 13.6 | 63 | 3,250 | 8-5/8 | 2.7 |
| 17 ..... | 27.3 | .737 | 26.1 | .152 | 13.3 | 79 | 3,250 | 8-5/8 | 4.0 |
| 18 ..... | 28.3 | .793 | 26.1 | .039 | 10.2 | 95 | 3,250 | 8-5/8 | 5.4 |
| 19 ..... | 44.6 | .777 | 42.8 | .236 | 15.1 | 70 | 3,250 | 8-5/8 | 4.4 |
| 20 ..... | 59.0 | .760 | 42.8 | .034 | 2.8 | 97 | 3,250 | 8-5/8 | 6.0 |
| 21 ..... | — | .786 | 23.1 | .032 | — | — | 4,000 | 8-5/8 | 5.6 |
| 22 ..... | 23.0 | .760 | 17.6 | .023 | 3.2 | 98 | 4,000 | 8-5/8 | 8.0 |
| 23 ..... | 25.4 | .750 | 24.0 | .078 | 12.8 | 90 | 4,000 | 8-5/8 | 4.0 |
| 24 ..... | 44.0 | .751 | 42.8 | .349 | 14.7 | 55 | 4,000 | 8-5/8 | 4.0 |
| 25 ..... | 44.5 | .745 | 42.8 | .165 | 12.1 | 78 | 4,000 | 8-5/8 | 5.6 |
| 26 ..... | 40.4 | .701 | 27.3 | .030 | 2.1 | 97 | 4,000 | 8-5/8 | 7.1 |
| 27 ..... | — | .487 | 42.8 | .040 | 1.2 | 98 | 4,000 | 8-5/8 | 7.1 |
| 28 ..... | 63.2 | .725 | 42.8 | .061 | 2.1 | 94 | 4,000 | 8-5/8 | 6.1 |

Table 5–13.—Example sludge thickening alternatives

| Alternative | Sludge thickening method: Primary | Waste activated | Combined |
|---|---|---|---|
| Number 1 ........................ | Gravity | Gravity | — |
| Number 2 ........................ | — | — | Gravity |
| Number 3 ........................ | None | DAF | — |
| Number 4 ........................ | Gravity | DAF | — |
| Number 5 ........................ | None | Centrifuge | — |
| Number 6 ........................ | Gravity | Centrifuge | — |

Required thickener: 9,208/8 lb/ft$^2$/day = 1,151 ft$^2$ (106.9 m$^2$) or two 27.1 ft (8.26 m) dia. units
Recommended thickener: two 30 ft (9.14 m) dia., 10 ft (3.05 m) deep units
Thickened product: 9,208/(0.05 × 8.34) = 22,082 gals/day (83,590 l/day)
Thickener cost: $160,000

The analysis has shown that capital costs for alternative No. 2 are slightly less than those for alternative No. 1 ($160,000 versus $162,000). Additionally, a thicker sludge would be obtained with alternative No. 2 (5 percent versus 4.6 percent). This would result in additional cost savings in the digestion facilities. A similar analysis for the 40 million gallons (1.75 m$^3$/s) per day plant resulted in a $167,000 unit (60-foot (18.29 m) diameter) for the primary sludge, and a $489,000 unit (110-foot (33.53 m) diameter) for the waste activated sludge (total cost $656,000), or two $305,000 units (85-foot (25.91 m) diameter) for the combined sludge (total cost $610,000). Thus, on the basis of capital costs, alternative No. 1 is deleted from further consideration.

Alternative No. 6 appears to be a viable solution for our example plant. However, an initial check of the thickened sludge product should be made since a sludge that is too concentrated can actually cause more problems in the anaerobic digestion facilities than a sludge which is too thin.

### Alternative No. 6

Definition—gravity thicken primary sludge; thicken waste activated sludge by centrifugation. Three shifts (24 hours) 7 day per week operation of gravity thickeners at both plants and of centrifuges at 40 Mgal/d (1.75 $m^3/s$) plant; two shifts (15 hours) 5 day per week operation of centrifuges at 4 Mgal/d (.18 $m^3/s$) plant.

Primary sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 5,204 (2360 kg/day) | 52,040 (23,605 kg/day) |
| Volume (gals/day) | 15,600 (59,050 l/day) | 156,000 (590,500 l/day) |

Recommended thickener
4 Mgal/d—one 20 ft (6.10 m) dia., 10 ft (3.05 m) deep unit
40 Mgal/d—one 60 ft (18.29 m) dia., 12 ft (3.66 m) deep unit
Thickened product
4 Mgal/d—6,933 gals/day (26,240 l/day) of 9 percent sludge
40 Mgal/d—69,330 gals/day (262,400 l/day) of 9 percent sludge
Thickener cost
4 Mgal/d—$64,000
40 Mgal/d—$167,000

Waste activated sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 4,004 (1816 kg/day) | 40,040 (18,161 kg/day) |
| Volume (gals/day) | 48,010 (181,740 l/day) | 480,100 (1,817,400 l/day) |

Recommended thickener
4 Mgal/d—one 75 gpm unit (4.73 l/s)
40 Mgal/d—one 667 gpm unit (42.08 l/s)
Thickened product (daily average based on 7 day week)
4 Mgal/d—8,002 gals/day (30,290 l/day) of 6 percent sludge
40 Mgal/d—80,020 gals/day (302,900 l/day) of 6 percent sludge
Thickener cost (based on one unit)
4 Mgal/d—$89,000
40 Mgal/d—$280,000

Combined product

4 Mgal/d—$[(6{,}933 \times 9) + (8{,}002 \times 6)]/(6{,}933 + 8{,}002) = 7.39$
14,935 gals/day (56,535 l/day) of 7.39 percent sludge
40 Mgal/d—149,350 gals/day (565,360 l/day) of 7.39 percent sludge

The calculations show that a 7.4 percent solids sludge would result. This exceeds the 4 to 6 percent solids recommended for efficient digester operation. Thus, alternative No. 6 is eliminated from further consideration. Detailed cost analyses are required for screening the remaining alternatives.

## Cost-Effectiveness Analysis

Design of the thickener units (based on data previously presented in this paper) and capital costs for those units will be presented first for the remaining alternatives. Other costs will then be analyzed.

### Alternative No. 3

Definition—thicken waste activated sludge with dissolved air flotation; no thickening of primary sludge; two shifts (15 hours) 5 days per week operation of DAF units at 4 Mgal/d (.18 $m^3/s$) plant; three shifts (24 hours) 7 days per week operation of units at 40 Mgal/d (1.75 $m^3/s$) plant.

Waste activated sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 4,004 (1816 kg/day) | 40,040 (18,161 kg/day) |
| Volume (gals/day) | 48,010 (181,740 l/day) | 480,100 (l,817,400 l/day) |

Required DAF equipment
4 Mgal/d—$(4{,}004 \times 7)/(15 \times 5 \times 2.0\ lb/ft^2/hr) = 187\ ft^2\ (17.4\ m^2)$
40 Mgal/d—$40{,}040/(24 \times 2.0\ lb/ft^2/hr) = 834\ ft^2\ (77.5\ m^2)$
Recommended DAF equipment
4 Mgal/d: two 100 $ft^2$ units (9.3 $m^2$)
40 Mgal/d: two 400 $ft^2$ units (37.2 $m^2$)
Thickened product (daily average based on 7-day week)
4 Mgal/d—$(4{,}004/0.04 \times 8.34) = 12{,}002$ gals/day (45,430 l/day)
40 Mgal/d—120,020 gals/day (454,330 l/day)
Thickener cost
4 Mgal/d—$82,000
40 Mgal/d—$205,000

Combined product (unthickened primary + thickened WAS)

4 Mgal/d—$15{,}600 + 12{,}002 = 27{,}602$ gals/day (104,490 l/day) of 4 percent sludge
40 Mgal/d—276,020 gals/day (1,044,900 l/day) of 4 percent sludge

### Alternative No. 4

Definition—gravity thicken primary sludge; thicken waste activated sludge with dissolved air flotation. Three shifts (24 hours) 7 days per week operation of gravity thickener at both plants and of DAF unit at 40 Mgal/d (1.75 $m^3/s$) plant; two shifts (15 hours) 5 days per week operation of DAF unit at 4 Mgal/d (.18 $m^3/s$) plant.

Primary sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 5,204 (2360 kg/day) | 52,040 (23,605 kg/day) |
| Volume (gals/day) | 15,600 (59,050 l/day) | 156,000 (590,500 l/day) |

Recommended thickener
4 Mgal/d—one 20 ft (6.10 m) dia., 10 ft (3.05 m) deep unit
40 Mgal/d—one 60 ft (18.29 m) dia., 12 ft (3.66 m) deep unit
Thickened product
4 Mgal/d—6,933 gals/day (26,240 l/day) of 9 percent sludge
40 Mgal/d—69,330 gals/day (262,400 l/day) of 9 percent sludge
Thickener cost
4 Mgal/d—$64,000
40 Mgal/d—$167,000

Final sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 4,004 (1816 kg/day) | 40,040 (18,161 kg/day) |
| Volume (gals/day) | 48,010 (181,740 l/day) | 480,100 (1,817,400 l/day) |

Recommended thickener
4 Mgal/d—one 200 $ft^2$ unit (18.6 $m^2$)
40 Mgal/d—one 800 $ft^2$ unit (74.3 $m^2$)
Thickened product (daily average based on 7-day week)
4 Mgal/d—12,002 gals/day (45,430 l/day) of 4 percent sludge
40 Mgal/d—120,020 gals/day (454,300 l/day) of 4 percent sludge
Thickener cost
4 Mgal/d—$55,000
40 Mgal/d—$91,000 (built-in-place unit, equipment only)

Combined product

4 Mgal/d—$[(6{,}933 \times 9) + (12{,}002 \times 4)]/(6{,}933 + 12{,}002) = 5.83$
18,935 gals/day (71,680 l/day) of 5.83 percent sludge
40 Mgal/d—189,350 gals/day (716,800 l/day) of 5.83 percent sludge

### Alternative No. 5

Definition—thicken waste activated sludge by centrifugation; no thickening of primary sludge. Two shifts (15 hours) 5 days per week operation of centrifuge units at 4 Mgal/d (.18 $m^3/s$) plant; three shifts (24 hours) 7 days per week operation of units at 40 Mgal/d (1.75 $m^3/s$) plant.

Table 5–14.—Thickener product and anaerobic digester requirements

| Alternative | Digester influent sludge | | | Digester volume ($ft^3$) | | Digester cost | |
|---|---|---|---|---|---|---|---|
| | Percent solids | Volume (gals/day) | | | | | |
| | | 4 Mgal/d | 40 Mgal/d | 4 Mgal/d | 40 Mgal/d | 4 Mgal/d | 40 Mgal/d |
| Number 2.................. | 5.0 | 22,082 | 220,820 | 58,034 | 580,340 | $789,000 | $4,074,000 |
| Number 3.................. | 4.0 | 27,602 | 276,020 | 71,938 | 719,380 | 877,000 | 5,310,000 |
| Number 4.................. | 5.83 | 18,935 | 189,350 | 49,683 | 496,830 | 742,000 | 3,425,000 |
| Number 5.................. | 4.68 | 23,602 | 236,020 | 61,661 | 616,610 | 806,000 | 4,361,000 |

Notes: If thickeners were not used, digester influent sludges would be as follows:
Alternative No. 2— 4 Mgal/d, 36,803 gals/day of 3.0%
40 Mgal/d, 368,030 gals/day of 3.0%
All other alternatives— 4 Mgal/d, 63,610 gals/day of 1.74%
40 Mgal/d, 636,100 gals/day of 1.74%

Digester design is based on the thickened sludge, 85°F. temperature, 20 days detention, 75 percent sludge volatile content, 2,302 pounds of dry solids per million gallons of wastewater and the volatile sludge loading factor method. Digester costs are based on two high rate units for each plant.

### Final sludge

| | *4 Mgal/d (.18 $m^3/s$)* | *40 Mgal/d (1.75 $m^3/s$)* |
|---|---|---|
| Quantity (lbs/day) | 4,004 (1816 kg/day) | 40,040 (18,161 kg/day) |
| Volume (gals/day) | 48,010 (181,740 l/day) | 480,100 (1,817,400 l/day) |

Recommended thickener
4 Mgal/d—two 38 gpm units (2.40 l/s)
40 Mgal/d—two 334 gpm units (21.07 l/s)
Thickened product (daily average based on 7 day week)
4 Mgal/d—4,004/(0.06 × 8.34) = 8,002 gals/day (30,291 l/day) of 6 percent sludge
40 Mgal/d—80,020 gals/day (302,910 l/day) of 6 percent sludge
Thickener cost
4 Mgal/d—$116,000
40 Mgal/d—$324,000

### Combined product

4 Mgal/d—[(15,600 × 4) + (8,002 × 6)]/(15,600 + 8,002) = 4.68
23,602 gals/day (89.340 l/day) of 4.68 percent sludge
40 Mgal/d—236,020 gals/day (893,440 l/day) of 4.68 percent sludge

The design calculations for the various alternatives indicate that they will result in different sludge moisture contents and sludge volumes. These data and the resultant required anaerobic digester volumes and costs are summarized in table 5–14. As shown by the data, considerable digester cost savings are possible with the thicker sludges.

The example plants are located in the Midwest. Therefore, the problem of possible freezing temperatures needs to be addressed. Except for icing of weirs and possibly a thinner product sludge, exposed gravity thickener operation should not be seriously affected in freezing weather. Flotation and centrifuge equipment, however, should be located in heated enclosures to prevent freezing of the exposed piping and to protect corrodible components from the elements. Besides housing the thickening equipment, the structure should also provide space for polymer feed equipment, and for polymer storage if polymers are to be used. At the example plants, assume that existing building space is fully utilized and, thus, any thickener building would be new construction. The required building space and associated costs for alternatives utilizing flotation or centrifugal thickening are presented in table 5–15. Polymers are required with alternatives Nos. 4 and 5. Storage space for a 30-day supply has been included in the required building area.

All capital costs for the alternatives have been summarized in table 5–16.

Power requirements and associated costs vary with the type and size of thickeners utilized. Gravity thickening systems require power for the operation of raw and

Table 5–15.—Required thickener building space

| Alternative | Thickener description | | Building description | | |
|---|---|---|---|---|---|
| | Type | Unit size | Area ($ft^2$) | Height (ft) | Building cost |
| #3–4 Mgal/d.... | DAF | 2–100 $ft^2$ | 1,520 | 12 | $84,000 |
| #3–40 Mgal/d.. | DAF | 2–400 $ft^2$ | 2,750 | 14 | 136,000 |
| #4–4 Mgal/d.... | DAF | 1–200 $ft^2$ | 1,150 | 14 | 75,000 |
| #4–40 Mgal/d.. | DAF | 1–800 $ft^2$ | 2,050 | 10 | [a]181,000 |
| #5–4 Mgal/d.... | Centrifugal | 2–38 gpm | 770 | 10 | 49,000 |
| #5–40 Mgal/d.. | Centrifugal | 2–167 gpm | 1,000 | 10 | 58,000 |

[a]Includes concrete tankage.

Table 5–16.—Capital costs

| Alternative description (Mgal/d) | Thickeners | Supportive equipment | Building | Anaerobic digester | Total |
|---|---|---|---|---|---|
| | | 4 Mgal/d plant | | | |
| Number 2 | $160,000 | $18,000 | — | $789,000 | $967,000 |
| Number 3 | 82,000 | 28,000 | $84,000 | 877,000 | 1,071,000 |
| Number 4 | 119,000 | 46,000 | 75,000 | 742,000 | 982,000 |
| Number 5 | 116,000 | 28,000 | 49,000 | 806,000 | 999,000 |
| | | 40 Mgal/d plant | | | |
| Number 2 | 610,000 | 24,000 | — | 4,074,000 | 4,708,000 |
| Number 3 | 205,000 | 44,000 | 136,000 | 5,310,000 | 5,695,000 |
| Number 4 | 258,000 | 68,000 | 181,000 | 3,425,000 | 3,932,000 |
| Number 5 | 324,000 | 44,000 | 58,000 | 4,361,000 | 4,787,000 |

thickened sludge pumps and the sludge collector drive. Dissolved air flotation systems also require power for raw and thickened sludge pumps, but additionally for a recirculation pump, reaeration pump (if present), bottom collector drive, skimmer drive, air compressor, polymer feed system (if present), and heating and lighting of the thickener building. Centrifugal thickening systems require power for the raw thickened sludge pumps, bowl drive, conveyor drive (if present), polymer feed system (if present), and heating and lighting of the thickener building. Since the required anaerobic digester volume differs with the four alternatives, the yearly sludge heating requirements will also vary. These sludge heating costs need to be included in the thickener cost-effectiveness analysis since they are directly related to thickening process. Total operating horsepower, thickener building heating requirements, and associated power costs for the various alternatives, excluding digester heating costs, are presented in table 5–17. Building lighting costs were determined insignificant and are not presented. Operating horsepower figures include influent and effluent sludge pump motors which total as follows: Alternate No. 2: 4 million gallons per day (.18 $m^3/s$)—1 horsepower (.75 kW), 40 million gallons per day (1.75 $m^3/s$)—5 horsepower (3.73 kW); Alternate No. 3: 4 million gallons per day (.18 $m^3/s$)—1-1/2 horsepower (1.12 kW), 40 million gallons per day (1.75 $m^3/s$)—4-1/2 horsepower (3.36 kW); Alternate No. 4: 4 million gallons per day (.18

Table 5–17.—Thickening power requirements and costs

<table>
<tr><th rowspan="2">Alternative (Mgal/d)</th><th colspan="3">Power requirements</th><th colspan="5">Yearly power costs</th></tr>
<tr><th colspan="2">Equipment (operating hp)</th><th>Heating (Btu/year)</th><th colspan="2">Equipment</th><th>Heating</th><th colspan="2">Total</th></tr>
<tr><td colspan="9">4 Mgal/d plant</td></tr>
<tr><td>Number 2, gravity thickener</td><td>5</td><td></td><td>—</td><td>$1,306</td><td></td><td>—</td><td>$1,306</td><td></td></tr>
<tr><td>Number 3, DAF thickener</td><td>50</td><td></td><td>$1.85 \times 10^8$</td><td>5,817</td><td></td><td>$765</td><td>6,582</td><td></td></tr>
<tr><td>Number 4, gravity thickener</td><td>2.5</td><td rowspan="2">42.5</td><td>—</td><td>653</td><td rowspan="2">5,306</td><td>—</td><td>653</td><td rowspan="2">5,981</td></tr>
<tr><td>Number 4, DAF thickener</td><td>40</td><td>$1.63 \times 10^8$</td><td>4,653</td><td>675</td><td>5,328</td></tr>
<tr><td>Number 5, centrifugal thickener</td><td>42.5</td><td></td><td>$8.60 \times 10^7$</td><td>4,944</td><td></td><td>355</td><td>5,299</td><td></td></tr>
<tr><td colspan="9">40 Mgal/d plant</td></tr>
<tr><td>Number 2, gravity thickener</td><td>11</td><td></td><td>—</td><td>2,874</td><td></td><td>—</td><td>2,874</td><td></td></tr>
<tr><td>Number 3, DAF thickener</td><td>140</td><td></td><td>$3.91 \times 10^8$</td><td>36,581</td><td></td><td>1,620</td><td>38,201</td><td></td></tr>
<tr><td>Number 4, gravity thickener</td><td>4</td><td rowspan="2">114</td><td>—</td><td>1,045</td><td rowspan="2">29,788</td><td>—</td><td>1,045</td><td rowspan="2">30,643</td></tr>
<tr><td>Number 4, DAF thickener</td><td>110</td><td>$2.08 \times 10^8$</td><td>28,743</td><td>855</td><td>29,598</td></tr>
<tr><td>Number 5, centrifugal thickener</td><td>106</td><td></td><td>$1.12 \times 10^8$</td><td>27,697</td><td></td><td>465</td><td>28,162</td><td></td></tr>
</table>

$m^3/s$)—2-1/2 horsepower (1.87 kW), 40 million gallons per day (1.75 $m^3/s$)—6-1/2 horsepower (4.85 kW); Alternate No. 5: 4 million gallons per day (.18 $m^3/s$)—1-1/2 horsepower (1.12 kW), 40 million gallons per day (1.75 $m^3/s$)—4-1/2 horsepower (3.36 kW). Power costs for equipment operation are based on a rate of $0.04 per kilowatt-hour ($1.11/mJ). Power costs for heating the building are based on using fuel oil at a cost of $0.45 per gallon ($.12/l). The cost associated with heating the sludge in the anaerobic digesters and the total power costs for each alternative are presented in table 5–18. In developing heating costs for the digester, it was assumed that auxiliary fuel (fuel oil at a cost of $0.45 per gallon ($.12/l)) would be required 50 percent of the time.

Polymers are required for dissolved air flotation thickening and may be required for centrifugal thickening of the waste activated sludge. Polymer requirements quoted by the various equipment manufacturers vary considerably for the same type process. Average polymer requirements based on several submittals and data from existing installations and the associated costs are presented in table 5–19.

Labor associated with operating and maintaining the thickening equipment varies with the complexity of the process. The continuously operating gravity thickener requires a visual inspection only once a shift, whereas the more complex dissolved air flotation and centrifuge systems should be checked every 2 or 3 hours. The inspections should be visual checks on the product quality and also on the operating conditions of all system components. Additional time for startup and shutdown of either the dissolved air flotation or centrifuge systems must be included if they are not operated on a continuous 24-hour basis (Alternatives Nos. 3, 4, and 5 for the 4 Mgal/d (.18 $m^3/s$) plant). Startup and shutdown time

Table 5–18.—Digester heating costs and alternative total power costs

| Alternative (Mgal/d) | Digester heating (Btu/year) | Digester heating (Cost/year) | Table 17 power costs (cost/year) | Total yearly power costs |
|---|---|---|---|---|
| | 4 Mgal/d plant | | | |
| Number 2 | $1.6831 \times 10^9$ | $6,749 | $1,306 | $8,055 |
| Number 3 | $2.0820 \times 10^9$ | 8,615 | 6,582 | 15,197 |
| Number 4 | $1.4563 \times 10^9$ | 6,026 | 5,981 | 12,007 |
| Number 5 | $1.7875 \times 10^9$ | 7,397 | 5,299 | 12,696 |
| | 40 Mgal/d plant | | | |
| Number 2 | $1.5628 \times 10^{10}$ | 64,668 | 2,874 | 67,542 |
| Number 3 | $1.9415 \times 10^{10}$ | 80,338 | 38,201 | 118,539 |
| Number 4 | $1.3426 \times 10^{10}$ | 55,556 | 30,643 | 86,199 |
| Number 5 | $1.6851 \times 10^{10}$ | 69,729 | 28,162 | 97,891 |

Table 5–19.—Polymer requirements and costs

| Alternative | Polymer required (lb/ton of dry solids) | Polymer cost: Unit ($/lb) | Polymer cost: Yearly total |
|---|---|---|---|
| Number 3 and 4– 4 Mgal/d DAF) | 35 | 0.08 | $2,046 |
| Number 3 and 4–40 Mgal/d DAF) | 35 | 0.08 | 20,460 |
| Number 5– 4 Mgal/d centrifugal) | 6 | 1.80 | 7,892 |
| Number 5–40 Mgal/d centrifugal) | 6 | 1.80 | 78,920 |

probably amounts to a total of about 1 hour per day. Routine sampling and testing of the thickener influent sludge, effluent sludge, and supernatant is required for any type thickener. The tests involved are essentially the same regardless of thickener type or size. Testing must be done more frequently, however, on DAF and centrifuge systems than on gravity systems. Routine maintenance includes such things as lubricating equipment and daily washdown or cleanup operations. At least once a year, all thickeners should be dewatered, throroughly inspected, and repaired, as necessary. Painting of corrodible components will probably be necessary at 5-year intervals. Solid bowl centrifuge conveyors may have to be resurfaced or replaced after 5,000–10,000 hours use, depending upon the amount of grit in the sludge and conveyor construction. A summary of the yearly operation and maintenance time and the associated costs for each alternative are presented in table 5–20.

Maintenance materials costs were developed from information provided by equipment manufacturers and data from existing installations. The materials costs shown in table 5–20 are estimates and, hence, may not be indicative of the costs associated with any one particular manufacturer's equipment. These costs may be described as percentages of the thickener system capital costs as follows: gravity thickening, 0.3 percent for small installations and 0.2 percent for larger installations; dissolved air flotation, 1 percent for small installations and 0.9 percent for larger installations; centrifugation, 5.2 percent for small installations and 3.1 percent for larger installations.

Power, chemicals, and operation and maintenance yearly costs have been summarized in table 5–21. Since the power requirements for the gravity thickening alternative (alternative 2) are low and chemicals are not required, it has the lowest yearly operating cost of all the alternatives. Although the centrifugation alternative (alternative 5) has power costs similar to those of the dissolved air flotation alternative (alternative 4), the yearly operating cost is considerably higher due to much higher chemical and operation and maintenance costs.

The alternatives' total capital costs and total yearly costs previously derived in tables 5–16 and 5–21, respectively, are repeated in table 5–22. The data show

Table 5–20.—Operation and maintenance time and costs

| Alternative description | Operator's time (hrs/year)[a] | Operator's time ($/year)[b] | Maintainer's time (hrs/year)[a] | Maintainer's time ($/year)[b] | Material cost ($/year) | Total cost ($/year) | |
|---|---|---|---|---|---|---|---|
| | | | 4 Mgal/d plant | | | | |
| Number 2, gravity ....... | 483 | $2,415 | 252 | $1,260 | $535 | $4,210 | |
| Number 3, DAF .......... | 1,416 | 8,496 | 586 | 3,516 | 1,100 | 13,112 | |
| Number 4, DAF .......... | 868 | 5,208 | 293 | 1,758 | 830 | 7,796 | 10,536 |
| Number 4, gravity ....... | 373 | 1,865 | 126 | 630 | 245 | 2,740 | |
| Number 5, centrifugal ... | 1,659 | 9,954 | 264 | 1,584 | [c]6,000 | 17,538 | |
| | | | 40 Mgal/d plant | | | | |
| Number 2, gravity ....... | 483 | 2,415 | 440 | 2,200 | 1,260 | 5,875 | |
| Number 3, DAF .......... | 2,496 | 14,976 | 804 | 4,824 | 2,240 | 22,040 | |
| Number 4, DAF .......... | 1,408 | 8,448 | 402 | 2,412 | 1,215 | 12,075 | 15,420 |
| Number 4, gravity ....... | 373 | 1,865 | 220 | 1,100 | 380 | 3,345 | |
| Number 5, centrifugal ... | 2,739 | 16,434 | 445 | 2,670 | [c]10,000 | 29,104 | |

[a]Time variances are due to equipment and operating time differences noted in the alternative definitions.
[b]Costs are based on $5/hr wage for gravity thickener operators/maintainers; $6/hr wage for DAF or centrifuge operators/maintainers.
[c]Costs are based on replacing conveyor after 7,500 operating hours.

Table 5–21.—Yearly operating cost summary

| Alternative description (Mgal/d) | Power | Chemicals | Operation and maintenance | Total |
|---|---|---|---|---|
| | 4 Mgal/d plant | | | |
| Number 2 ............... | $8,055 | — | $4,210 | $12,265 |
| Number 3 ............... | 15,197 | $2,046 | 13,122 | 30,355 |
| Number 4 ............... | 12,007 | 2,046 | 10,536 | 24,589 |
| Number 5 ............... | 12,696 | 7,892 | 17,538 | 38,126 |
| | 40 Mgal/d plant | | | |
| Number 2 ............... | 67,542 | — | 5,875 | 73,417 |
| Number 3 ............... | 118,539 | 20,460 | 22,040 | 161,039 |
| Number 4 ............... | 86,199 | 20,460 | 15,420 | 122,079 |
| Number 5 ............... | 97,891 | 78,920 | 29,104 | 205,915 |

Table 5–22.—Cost summary and rank

| Alternative description | Capital costs | Ranking | Yearly operating costs | Ranking |
|---|---|---|---|---|
| | 4 Mgal/d plant | | | |
| Number 2 .................. | $967,000 | 1 | $12,265 | 1 |
| Number 3 .................. | 1,071,000 | 4 | 30,355 | 3 |
| Number 4 .................. | 982,000 | 2 | 24,589 | 2 |
| Number 5 .................. | 999,000 | 3 | 38,126 | 4 |
| | 40 Mgal/d plant | | | |
| Number 2 .................. | 4,708,000 | 2 | 73,417 | 1 |
| Number 3 .................. | 5,695,000 | 4 | 161,039 | 3 |
| Number 4 .................. | 3,932,000 | 1 | 122,079 | 2 |
| Number 5 .................. | 4,787,000 | 3 | 205,915 | 4 |

that for the 4 million gallons per day (.18 $m^3/s$) plant, the least expensive option in terms of both capital and operating costs is gravity thickening of the combined sludge followed by anaerobic digestion (alternative 2). Note that there is only 3 percent difference between the capital cost of alternative 2 and the third most expensive alternative (in terms of capital costs—alternative 5).

The results for the 40 Mgal/d (1.75 $m^3/s$) plant are somewhat different than those for the 4 Mgal/d (.18 $m^3/s$) plant. In this case, the least costly alternative in terms of capital costs does not correspond with the least costly one in terms of yearly operating costs. Additionally, for the 40 Mgal/d (1.75 $m^3/s$) plant, the least costly alternative (capital costs) is not alternative No. 2 (as was the case for the 4 Mgal/d (.18 $m^3/s$) plant) but alternative No. 4. Also, in this case, there is a 22 percent difference between the capital cost of the least expensive and the third most expensive alternative. Since the lowest capital cost and lowest operating cost alternatives do not correspond, a present worth analysis

Table 5–23.—Present worth analysis 40 Mgal/d alternatives No. 2 and No. 4

| | Alternative No. 2 | Alternative No. 4 |
|---|---|---|
| Construction cost | $4,708,000 | $3,932,000 |
| Project cost[a] | 5,273,000 | 4,482,500 |
| Project contingency | 527,300 | 448,300 |
| Total cost for facilities | 5,800,300 | 4,930,800 |
| Constant I-20 O. & M. costs | 73,400 | 122,100 |
| Salvage value | 1,527,200 | 1,247,000 |
| Present worth | | |
| Initial project | 5,800,300 | 4,930,800 |
| P.W. constant O. & M. | 800,800 | 1,332,100 |
| Subtotal | 6,601,100 | 6,262,900 |
| P.W. salvage value | 423,300 | 345,700 |
| Total present worth | 6,177,800 | 5,917,200 |
| Average annual equivalent cost | 566,500 | 542,600 |

[a]Includes costs associated with engineering, legal and administrative, inspection, surveying, soil borings, start-up and generation and maintenance manual, and interest during construction.

would be required to make the final selection. Although the alternative capital cost rankings varied with plant capacity, the yearly operating cost rankings did not. Gravity thickening of the combined sludge followed by anaerobic digestion had the lowest operating costs; centrifugation of the waste activated sludge or thickening of primary sludge, followed by anaerobic digestion, had the highest. A present worth analysis is presented in table 5–23. This analysis shows that alternative No. 4 (gravity thickened primary sludge and DAF thickened waste activated sludge) has the lowest average annual equivalent cost for the 40 Mgal/d (1.75 $m^3/s$) facility.

## SUMMARY

The purpose of this paper has been to describe, in detail, those thickening methods which are currently utilized, and to present the general approach necessary in evaluation of thickening alternatives by means of a design example. The methods presented can be used to analyze a thickening problem at any wastewater treatment plant, regardless of its size or complexity. The results of the design example are valid for the assumptions made. Any change in problem definition could mean a different solution.

Recommendation of a particular process should be geared to available operation and maintenance personnel. Considerably more skill is required to operate and maintain dissolved air flotation and centrifuge equipment than gravity thickeners. The final recommended alternative process will be one that is agreed upon by the owner, the engineer, and the regulatory agency.

## REFERENCES


1. USEPA, "Process Design Manual for Upgrading Existing Wastewater Treatment Plants," USEPA Technology Transfer EPA-625/1-74-004a, October 1974.
2. USEPA, "Process Design Manual for Sludge Treatment and Disposal," USEPA Technology Transfer, EPA-625/1-74-008, October 1974.
3. Water Pollution Control Federation, "Operation of Wastewater Treatment Plants,"—Manual of Practice No. 11, WPCF, Washington, D.C., 1976.
4. Metcalf and Eddy, Inc., "Wastewater Engineering," McGraw-Hill, New York, 1972.

# Review of Developments in Dewatering Wastewater Sludges

## INTRODUCTION

This chapter reviews the sludge dewatering operating experiences which have occurred over the past 4 to 6 years and assesses the impact of these results on future designs. Particular emphasis is placed on innovative concepts and equipment.

It should be noted that practically all of the innovative development of new dewatering equipment in the last 4 to 6 years occurred first in Europe (particularly in the Federal Republic of Germany) and Japan, and has only begun to be transferred and utilized in the United States very recently. It will also become apparent that older, previously dominant equipment and concepts have, in many instances, been replaced as a direct consequence of operating results.

In attempting to incorporate the latest and best equipment and concepts into current design, the U.S. design engineer must be aware of plant operational results with various alternate systems.

Alternative dewatering equipment and procedures cannot be evaluated in isolation, but only as part of an overall system conceptual design. The interrelationship between sludge processing and liquid stream processing should always be considered. Previous works[1,2] graphically illustrate the adverse effects of equipment or systems that provide less than 90 percent capture of influent solids, and thereby illustrating the profound effect of the choice of a dewatering system on the operability and cost-effectiveness of the liquid processing system. Events of the past 4 to 6 years have further verified this principle.

There have been indications that the selection of the type of activated sludge system could have a strong effect on the relative severity of associated sludge processing problems.

Table 6–1 lists the effect of various activated sludge process modifications on yields of excess biomass and on sludge processability.

While this table is a summary estimate, the trends and principles involved are accurate. Given the above information, it is understandable why some states have banned the "High Rate" version of the activated sludge process. Regardless of statutory positions, results at plants incorporating "High Rate Activated Sludge" are sufficient to deter its use if the resultant sludge is to be disposed of in other than liquid form.

In selecting a dewatering system, an item of real concern is the choice of the final or ultimate disposal method for the sludge or its residue. Indeed, the available options for final disposal should be known prior to selection of the dewatering system. Fortunately, some of the new dewatering equipment, by virtue of producing higher solids content dewatered cake and by offering the capability to eliminate inorganic conditioning solids in dewatered cakes, provides considerably more flexibility than was previously available in matching up a dewatering process and an ultimate disposal process.

Dewatering is essentially always preceded by thickening and conditioning, and frequently by stabilization. The essential role of dewatering is to transform a dilute water slurry into a damp, moist cake form for either direct final disposal or for drying as a final product, or for reduction via an incineration or other combustion process prior to final disposal.

In evaluating dewatering processes it is essential to

Table 6–1.—Excess biomass production and sludge processability from various activated sludge processes

| Process variation | Pounds biochemical oxygen demand per 1,000 $ft^3$ | Food to microorganism ratio | Pounds W.A.S. pound biochemical oxygen demand removed (typical) | Sludge processability |
|---|---|---|---|---|
| High rate | 100–1,000 | 0.4–1.5 | 1.07 | Poor |
| Conventional | 20–40 | 0.2–0.4 | 0.4 | Good |
| Extended aeration | 10–25 | 0.05–0.15 | 0.15 | Variable |

Table 6–2.—Autothermic combustion[4]

| Parameter | Case A | Case B |
|---|---|---|
| Gross calorific value | 17,400 | 29,100 |
| Percent combustible matter in solids | 60 | 75 |
| Percent solids for autothermicity | 41.8 | 18.5 |

consider more than the direct operating costs, the production rate, and the dry solids content of the dewatered cake. The evaluation should include complete material balances (Quantified Flow Diagram 1, 2) around the dewatering process with a concurrent evaluation of the effect of all recirculation streams on other preceding unit processes, and the effect of all dewatered cake properties on the processes subsequent to dewatering, including final disposal.

To illustrate this point, note in table 6–2 that the percent dry solids level at which autogenous incineration occurs is a function of the calorific value of dry solids in dewatered cake, which in turn varies with the chemical composition of the solids. The requisite dry solids level for self sustaining combustion varies from 18.5 to 41.8 percent depending on these factors which are in turn materially affected by the unit processes to which the sludge has been subjected prior to dewatering.

## ANALYSIS OF RECENT PLANT OPERATING RESULTS AND IMPLICATIONS FOR DESIGN

What lessons should the past 5 years of plant operating results bring to bear on current and future designs? The following five points which bear consideration:

1. The effect of choice of type of biological process on sludge processing, and vice versa.
2. The effect of the inclusion of biomass on the sludge processing system.
3. The effects of processing discontinuity on biomass or mixed sludge processability.
4. The importance of painstaking analysis of plant results.
5. Relative operability and maintainability of various sludge processing systems or units.

### Type of Biological Process Chosen

As previously noted in table 6–1, the selection of the High Rate activated sludge process variation can result in a plant having to process a mixed sludge with 65 percent or greater biomass content. Further, that particular type of biomass is normally much more difficult to process than other types. While imposition of other design constraints may have resulted in utilization of the High Rate process in certain cases, it is apparent that a current overall system evaluation of alternate conceptual designs, particularly in the light of operating experiences, would usually not support the use of the High Rate system.

Results have also shown that the extended aeration process, unless kept within certain food to microorganism (F/M) and solids retention time (SRT) ranges can cause sludge processing problems. These factors further strengthen the need for adequate testing of sludges from alternate biological processes prior to selection of same.

### Effects of Inclusion of Biomass

The results of the past 5 years are reflected in the following list:

1. Gravity thickening of mixtures of primary and excess biomass sludges is usually ineffective (unless flocculants are used).
2. Recycling of biomass to primary clarifiers is nearly always a self-defeating process which causes more problems than it cures.
3. Inclusion of biomass in a mixture with primary sludge causes settling problems in conventional two-stage anaerobic digestion systems. This, plus the need to maximize gas production frequently makes single stage complete mix anaerobic digestion the process of choice for stabilization prior to dewatering in plants where sludge stabilization is required prior to dewatering.
4. Biomass causes poor settleability in elutriation tanks. These tanks can be modified to serve as post digestion thickening tanks (with use of flocculants). This is essential for economic dewatering.
5. Inclusion of biomass makes the careful selection of dewatering systems, including pretreatment processes such as conditioning and thickening, essential to successful design.

### Processing Discontinuity and Sludge Storage Effects

The following list delineates the pitfalls inherent in excessive accumulation of sludge within a plant brought on by discontinuity in removal by the dewatering process, either by excessive unplanned down time, or by design.

1. Development of septicity.
2. Destruction of some of the bioflocculation of the biomass.
3. Partial solubilization through prolonged aqueous contact.
4. Increased hydration and more sensitivity to shear (pumping, etc.).
5. Deterioration of processability occasioned by all four of the preceding.

### Methods of Analyzing Plant Operating Results

In considering the significance of plant results and relevance to design decisions, the following four concepts bear consideration:

1. The use of single static numbers as bench marks for a dynamic, interrelated system can be seriously misleading.
2. Appreciation of the "Inertia" inherent in moderate and large plant processing systems is necessary.
3. There is a paramount need to maintain "Steady State" conditions as much as possible.
4. Recycle or sidestreams should be minimized within reasonable ranges.

In developing design criteria, it is frequently assumed that dewatering equipment can be sized using steady state flow conditions for the overall system with some allowance for peaking. These assumptions are reasonable as long as reliable conditioning, thickening, and dewatering equipment are installed. However, if sludge removal operations are interrupted for lengthy periods or fundamental changes are made, then the standard factors may not be applicable in terms of order of magnitude. While some properly aerated sludge storage capacity is beneficial, storage usage should be minimized and septicity avoided whenever dewatering is used.

The length of time required to reestablish equilibrium or steady state conditions in moderate or large size plants with significant inventories of sludge is much longer than would normally be anticipated. This "Inertia Factor" is calculable through the use of mathematical models. From experience, in large plants, it can take several months to fully evaluate the effect of changes.

The need to maintain a "Steady State" or equilibrium removal rate of sludge sufficient to prevent overaccumulation within a plant is paramount. Once an accumulation problem develops, rapid resolution via accelerated removal rate procedures will prevent further difficulty.

Particular attention must be paid to processes which inherently cause heavy recycle loads. Processes or equipment which cause heavy recycle loads can have a negative effect on sludge removal rate. If large quantities of sludge have accumulated in a plant either because of heavy recycle loads or from a shut down period, normal operating schedules will require alteration. In order to clean out such an accumulation the "Sludge Removal Rate" during the transition "Clean Out" period prior to reestablishment of a normal equilibrium must be much greater than the normal rate. Unfortunately, if the overaccumulation is due to processes or equipment which cause a significant recirculation load of biomass, the aeration system of the plant will, during the "Clean Out" period of higher than normal sludge removal rate, be extremely overloaded and will also produce more excess activated sludge than normal. Another effect is that sludge storage renders biomass more difficult to process and results in a much greater amount of recirculation than normally would be predicted by "standard condition" testing figures and criteria.

### Relative Operability and Maintainability of Various Dewatering Systems and Units

The reliability and maintenance characteristics associated with various types of conditioning-dewatering processes, equipment, and brands is very important to the municipality and its personnel, and ultimately to the public who pays the bill. In addition to the need to keep units operating to prevent sludge accumulation and its attendant bad effects, maintenance costs are a very important factor in overall system costs.

The only truly accurate source of reliability and maintenance cost data is actual plant operational results. To justify professional process and equipment selection, the design engineer should acquaint himself thoroughly with reliability and maintenance parameters by visiting existing installations and obtaining accurate information from operating personnel. It is also necessary to sort out when problems are due to poor plant maintenance practices and when they are due to inherent process or equipment characteristics. If performance data are not available then they should be specified and a guarantee provided by the supplier.

The current methodology of bidding and selection of suppliers to equipment municipal plants has been, in some cases, a cause of some of the reliability and maintenance problems now being experienced. The bidding documents or plans and specifications should include cost factors for maintenance and life cycle, and should be sufficiently complete to ensure that truly equal equipment specified is provided. If this is not done, and the job is awarded on a strictly lowest price basis, inferior processes and equipment can be selected.

## CONDITIONING FOR DEWATERING

The following list delineates the normal functions of conditioning for dewatering:

1. Flocculation of suspended solids (particularly fines).
2. Washing out the alkalinity of anaerobically digested sludge (the original purpose of elutriation).
3. Promotion of rapid formation of a stable drainable cake.
4. Promotion of cake release from filtration support media.
5. Enhancement of cake fuel value.
6. Prevention of scale formation and corrosion inhibition.

The methods used to accomplish the above functions are as follows:

1. Chemical addition (inorganic).
2. Chemical addition (organic flocculants).
3. Elutriation (new function).
4. Heat treatment (conversion).

5. Ash addition (cake release).
6. Coal addition (fuel value).
7. Polyphosphonate addition (scale inhibition).

### Chemical Conditioning

In organic chemical conditioning the most notable occurrence has been the increase in total availability of metal salts, such as ferric chloride, due to the entry into the market of firms recovering the products from waste acids.

The dosage and costs for chemical conditioning vary substantially for activated sludge plants depending on the type of biological system employed and the overall sludge processing system. For this reason, unless otherwise noted, all quotations of typical dosage figures assume that plants are well designed, not involving procedures or systems which are known to materially increase conditioning demand and costs. An example of the latter would be to pipeline sludge for several miles prior to dewatering.

In the organic polyelectrolyte flocculant area, there have been several developments of consequence to dewatering processes:

1. New high charge density, high molecular weight materials in dry powder form, which are more efficient in conditioning the difficult sludges, have become available and are used widely.
2. A new class of compound, the "Mannich" cationic products, which have different performance characteristics, have been introduced, almost entirely as liquid products. These materials produce a floc and drainage characteristic more akin to that produced by ferric chloride.
3. Emulsion form cationic products of high charge density and molecular weight have been developed and are used.

### Elutriation

This process had been applied successfully to digested primary sludge, was misapplied to mixtures of primary and biomass sludges, and then adapted very successfully as a flocculant aided postdigestion thickening process to facilitate cost-effective dewatering.[1]

### Heat Treatment

This type process, sometimes called "Thermal Conditioning," is covered in detail in Chapter 4.

City utilities which have had dewatering experiences of note, some written up in the literature and some not, are:

- Kalamazoo, Mich.
- Colorado Springs, Colo.
- Chattanooga, Tenn.
- Chicago, Ill.
- Columbus, Ohio
- Perth, Scotland
- Ft. Lauderdale, Fla.
- Port Huron, Mich.
- Flint, Mich.
- Lakeview, Ontario
- Green Bay, Wis.

In Great Britain, where the most and earliest installations of the Porteous and Farrer heat treatment processes were made, the heat treatment process has been largely abandoned. In one case, a new plant, never used, has been offered for sale.

British water authorities detected significant quantities of refractory organic material in the effluent from plants dewatering heat treated sludges. The authorities consequently banned recycle of cooking liquors into biological treatment systems which discharge into rivers subsequently used as sources of drinking water, since the biological systems are incapable of removing the refractory organic material.

An additional development in dewatering heat treated sludges has been the need to chemically condition sludges in a number of cases, either on a spasmodic or regular basis. In the case of Port Huron, Mich. (Farrer System), which employs centrifuges for dewatering, routine use of flocculants at the rate of $8/ton ($8.82/Mg) of sludge dewatered has been found necessary. Other heat treatment plants have found flocculants necessary to promote cake formation to obtain reasonable solids capture.

To help alleviate scaling problems, Grand Rapids has found it necessary to condition heat treated sludge with $3/ton ($3.31/Mg) of polyphosphonates.

Various other chemicals have been found necessary to raise the pH of sludges, to condition boiler feed water, and to solvent wash scale from heat exchangers.

## DEWATERING EQUIPMENT TRENDS

The following is a list of the types of dewatering equipment or processes normally used in municipal wastewater sludge processing:

1. Drying beds.
2. Rotary vacuum filters.
3. Horizontal solid bowl centrifuges.
4. Pressure filters.
5. Continuous belt filter presses.
6. Rotating cylindrical devices.
7. Imperforate basket (batch) centrifuges.
8. Lagoons.

Drying beds are widely used at a large number of plants, particularly moderately sized plants in sunny climes, but not restricted to same. As will be seen, they have been the subject of recent developmental improvement activity, both with regard to improved capacities and mechanical removal facilities.

Whereas rotary vacuum filters were once the common of mechanical dewatering systems, their incidence of selection has rapidly decreased due to energy costs, the problem of cake pick up with certain sludges, and lack of ability to provide as dry a dewatered cake as various other devices.

Horizontal solid bowl centrifuges, particularly of the new low speed type, are still popular where a very high solids cake is not essential. Their popularity has dwindled to some extent due to energy considerations.

Pressure filters of the ordinary recessed chamber type have been installed in a few U.S. plants. Results have been mixed insofar as overall performance is concerned, despite the attainment of somewhat higher total cake solids levels (without necessarily improving the ratio of sewage solids to water) compared to Rotary Vacuum Filters or Solid Bowl Centrifuges. Major problems are cost, maintenance, and the frequent need to use high percentages of inorganic conditioners.

The new continuous belt filter presses have become the most widely selected dewatering devices for municipal sludge dewatering. Their rapid growth in popularity is due to ease of operation, low energy consumption, and the ability (in some models) to produce dewatered cakes with solids contents much greater than obtainable with Rotary Vacuum Filters, Centrifuges, or conventional Pressure Filters.

Rotating cylindrical devices, such as the Pernutit DCG, have been installed in some plants. Their use has been primarily at small plants and as the first stage of a dual system which includes an inclined multiroll press (MRP) for further cake dewatering.

Imperforate basket batch centrifuges have been installed at a few small plants where a low solids, relatively fluid cake is tolerable.

Lagoon drying is now frequently applied.

## DEWATERING METHODOLOGY

Wastewater sludges all form cakes during the dewatering process which are compressible to some degree and by virtue of this fact and their inherent water binding nature tend to require application of conditioning processes to facilitate a reasonable dewatering rate.

The various sludges may be indexed or characterized by determination of the "Specific Resistance to Filtration." They may also be characterized by being subjected to standardized bench scale dewatering test procedures (Filter leaf or Buchner funnel tests).

An important facet for design consideration is that dewatering of wastewater sludges is a "Cake Filtration" process. The cake which forms during dewatering is the primary filtration media and relative cake structure and form throughout the dewatering process will largely determine the efficacy of the system.

In assessing the cost-effectiveness of the pretreatment methods aimed at improving dewatering it is essential that the effect of these processes on the type of cake formed be considered. In most municipal wastewater treatment plants, if the following steps are effected, a mixed primary and biological sludge will result which is amenable to a cost-effective dewatering process yielding a dewatered cake suitable for either reduction or direct ultimate disposal in an economic fashion:

1. Maximization of solids capture in well-designed primary basins so as to provide as much typically easy to process "Primary" sludge as possible. This precludes high recycle loads of W.A.S. or thickener overflows or heat treat cooking liquors to the primary basins.
2. Selection of biological process variation with reasonable assessment of the amount and type of excess biomass which will be produced and will have to be processed. This usually precludes use of "High Rate Activated Sludge" processes and some Extended Aeration designs.
3. Use of gravity sludge thickeners only for straight primary sludge, or if this is not possible, provision of flocculant dosage capability to ensure reasonable solids capture and underflow thickened sludge solids contents when mixed primary-biological sludge is being thickened.
4. Use of dissolved air flotation or centrifugal thickening for excess activated sludge prior to mixed sludge anaerobic digestion, or prior to dewatering if stabilization is not to be included.
5. If anaerobic digestion of mixed sludge is employed, use of a single stage complete mix process and a post digestion thickening process, either gravity settling or dissolved air flotation (DAF).
6. Use of a conditioning process which does not result in creation of a heavy recycle load, either in the form of suspended or dissolved solids or in the form of $BOD_5$ or chemical oxygen demand (COD) or refractory organics. Likewise the conditioning process should not destroy any significant amount of the matrix forming material in the sludge solids which will form the cake in the dewatering process, and should not alter other cake properties requisite to the succeeding processes.
7. Selection and use of a dewatering device which is of rugged design, readily maintainable and will provide a minimum solids capture of 90 percent and a cake solids content amenable to succeeding processes. It is, for all practical purposes, always necessary to condition municipal sludges prior to dewatering.

## DRYING BEDS

Sludge drying beds are frequently referred to as "Sand Beds." In most cases except instances wherein "paved drying beds" or wedge water screens are used, sand is the primary drainage and cake support medium. The recent and continuing development of various types of Drying Beds prompts the use of that term, rather than Sand Bed.

Drying Beds are still the most common method of municipal wastewater sludge dewatering. The only reason they are not widespread in use is that they have not been the subject of any significant degree of development and improvement. This situation is changing as municipalities become more cognizant of their viability and relatively low cost of construction, operation, and maintenance when properly designed. An additional previous deterrent to their use has been the frequent lack of inclusion of mechanical sludge removal capability and an understandable dislike by operating personnel, occasioned by a need for manual removal. This deterrent can be and has been removed in many cases by relatively minor design modifications to facilitate mechanical removal.

An additional previous deterrent to selection of the drying bed alternative is that the "Ten State Standards" do not reflect the application of conditioning to sludges prior to dewatering. The use of "Ten State Standards" criteria, which assume no sludge conditioning, can result in excessive land requirements and the resultant acquisition costs artificially inflate cost estimates for the drying bed alternative.

A well-designed and properly operated drying bed can produce a drier sludge than any mechanical device. They are also less sensitive to the influent solids concentration.

On the negative side, drying beds are generally applicable only to digested or stabilized solids. Though they are particularly suitable for small installations and the "Sun Belt," drying beds are used successfully in treatment plants of all sizes and in widely varying climates (i.e., Chicago southwest treatment plant, the largest plant in the world).

Drying beds may be roughly categorized as follows:

1. Conventional rectangular beds with side walls, layers of sand and then gravel with under drainage piping to carry away the liquid. They are built either with or without provision for mechanical removal and with or without either a roof or a greenhouse type covering.
2. Paved rectangular drying beds with a center sand drainage strip with or without heating pipes buried in the paved section and with or without covering to prevent incursion of rain.
3. "Wedge-Water" drying beds which include a wedge wire septum incorporating provision for an initial flood with a thin layer of water, followed by introduction of liquid sludge on top of the water layer, controlled formation of cake, and provision for mechanical cleaning.
4. Rectangular vacuum assisted sand beds with provision for application of vacuum as a motive force to assist gravity drainage.

## Mechanism

On drying beds, the dewatering initially proceeds by drainage and then continues by evaporation. The proportion and absolute amount achieved by drainage will vary depending on whether or not the cake has been conditioned, and its overall drainage characteristics. An important consideration is the relative time period required for the cake to develop cracks which expose additional sludge to evaporation effects. Since one of the main functions of conditioning is to flocculate and immobilize the smaller "fines" particles in the sludge cake it is immediately apparent why a conditioned sludge slurry dewaters in a fraction of the time required for an unconditioned sludge. The completion of the drainage period is substantially delayed in an unconditioned sludge by migration of the fines to the sludge cake sand interface resulting in some plugging of the uppermost layer of sand. Maintenance of porous, relatively open structure within the cake is also essential to evaporation rate.

## Conventional Rectangular Beds

Drying bed drainage media normally consists approximately as follows:

1. The top layer is 6 to 9 inches (15.2 to 22.9 cm) of sand, usually with an effective size of 0.3 to 1.2 mm and a uniformity coefficient less than 5.
2. About 8 to 18 inches (20.3 to 45.7 cm) of gravel with size gradation of 1/8 to 1.0 inch (0.3 to 2.3 cm). The top three inches (7.6 cm) of the gravel layer are preferably 1/8 to 1/4 in. (0.3 to .6 cm) size.
3. Underdrain piping with a minimum diameter of 4 inches (10.2 cm) is often vitrified clay with open joints spaced 8 to 20 feet (2.44 to 6.10 cm) apart. Recently, plastic pipe is being used to prevent possible cracking when front end loaders are run across the bed for sludge removal. If a gridwork of concrete runways is provided for the front end loader, the selection of pipe is not critical.

Drying beds are frequently enclosed by glass. The glass enclosures can materially improve the performance of the beds, particularly in cold or wet climates. Experience has shown that in some cases only 67 percent of the area required for an open bed is required with enclosed beds. The degree to which, at specific locations, the space requirement could be reduced and the sludge loading increased by use of translucent roofing or total glass enclosure is a function of site rainfall, temperature, and sunlight prevalence.

Unfortunately, mechanical removal methods have not normally, in the past, been used with glass enclosed beds. Obviously the adaptation would not be either difficult or expensive.

Table 6–3 describes the typical design criteria for open drying beds.

The combination of the use of chemical conditioning plus design to permit mechanical sludge removal coupled with the use of either a translucent roof or complete glass enclosure with ventilation louvers dramatically lowers the space requirement for conventional drying bed use and should be the first alternative considered for dewatering in most plants.

The sidestream from drying bed operation consists of

Table 6–3.—Criteria for design of open conventional drying beds

| Type digested sludge | Pre-treatment | Area (sq.ft./cap.) | Sludge loading dry solids (lb/sq. ft./yr.) |
|---|---|---|---|
| Primary and humus...... | None | 1.6 | 22 |
| Primary and activated... | None | 3.0 | 15 |
| Primary and activated... | Chemically conditioned | 0.64 | 55 |

the drainage liquor which may be augmented by rainfall in the case of open beds. The additional drainage water is not normally a problem. The drainage water is usually relatively innocuous and can be recyled into the plant with impunity.

Drying times in open beds also vary due to climate, type of sludge, and whether or not it has been conditioned. In good weather, an average of 45 days is reasonable for unconditioned sludge. This period can be reduced to 5–15 days or less via conditioning.

A typical case study of the use of conventional drying beds follows.

## TAMPA, FLA.—CURRENT PLANT

The current Tampa plant is a primary treatment facility featuring anaerobic digestion and sand drying beds for sludge dewatering. The plant is designed for a flow of 36 Mgal/d (1.58 $m^3/s$) and is normally treating 40 Mgal/d (1.75 $m^3/s$). On occasion, alum and polyelectrolyte are used in the liquid treatment phase to meet the current interim effluent standards.

### Drying Bed Details and Operations—Existing Tampa Primary Plant

Thirty-three beds, each 125 by 60 feet (38.10 by 18.29 m) are employed. The rectangular beds employ a drainage medium of two sizes of graded sand above two layers of differently sized stone or gravel. The beds are usually refurbished every 2 to 3 years, at most. Current anaerobically digested primary sludge production is estimated to be 56,000 gallons (211,980 l) of 3.0 percent dry solids content per day. This is equivalent to 14,000 pounds/day (6350 kg/day) of dry solids. With 33 beds of 7,500 square feet (696.8 $m^2$) area each, the total available drying area is 247,500 square feet (22,993 $m^2$).

The 33 older drying beds at Tampa are not covered so the drying cycle varies somewhat due to rainfall variation. Nonetheless, the operation has been so successful that the new expanded advanced waste treatment (AWT) plant which will be in operation shortly is also equipped with drying beds for sludge dewatering. Tampa has for about 3 years regularly used polyelectrolytes for conditioning the sludge on its way into the drying beds. Drying time to liftable cake conditions without conditioning used to run 30 days minimum. With chemical conditioning, the drying time varies from 8 to 15 days depending on rainfall pattern.

Tampa features front end loader mechanical removal of dried sludge cake from the beds. One man can easily empty one bed in 6–8 hours. Previous removal methods involved use of 5 men for 1-1/2–2 days to remove sludge from one bed.

Figure 6–1 is a photograph of the mechanized sludge removal equipment used at Tampa on the drying beds.

Current operating procedure involves pumping about 55,000 to 60,000 gallons (208,200 to 227,130 l) of digested sludge onto a bed with in-line dosing of cationic liquid polyelectrolyte at a dosage rate of about 50 pounds per ton (25 kg/Mg). The price of the liquid cationic polymer is $0.13 per pound (0.29 per kg) on an as is, liquid basis making the conditioner cost $6.50 per ton ($7.17 per Mg) of dry solids.

Taking the estimated bed loading volume of 56,000 gallons (211,980 l) of 3.0 percent sludge and an average drying time of 11.5 days, the solids loading rate on the current Tampa beds is 60 pounds/square foot/year (292.8 $kg/m^2/year$). It should be noted that current practice is to produce a very dry cake as shown in figure 6–2.

## TAMPA, FLA.—NEW ADVANCED WASTE TREATMENT PLANT

Tampa has installed and is now starting up a new plant which features biological nitrification and denitrification with chemical addition for phosphorous removal. The

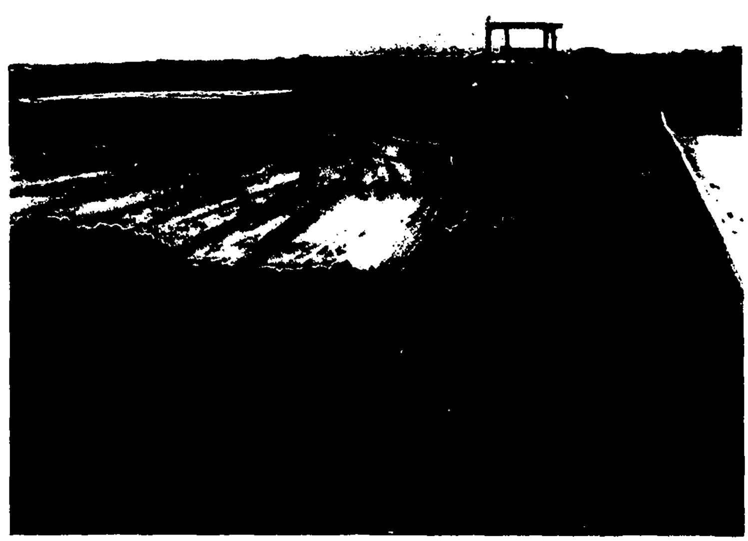

Figure 6–1.—Mechanized sludge removal at Tampa.

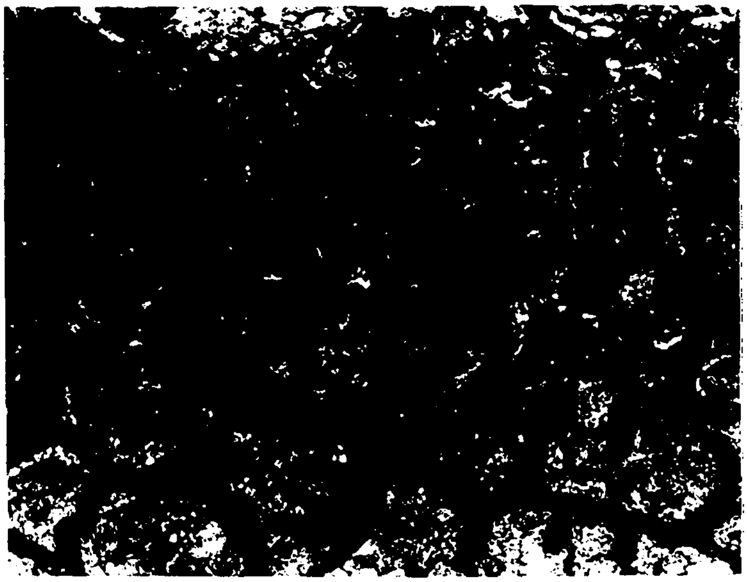

Figure 6–2.—Dried cake appearance on bed at Tampa before removal.

new plant is designed for a treatment capacity of 60 million gallons per day (2.63 $m^3/s$).

Aerobic digestion for advanced waste treatment plant (AWT) sludges and anaerobic digestion for primary sludge plus 32 new sand drying beds (each 100 by 140 feet 30.48 by 42.67 m) were included in the new facility.

While the original concept of the new facility was to aerobically digest the excess biological solids and to dewater them separately on the new drying beds, along with the AWT chemical solids, considerable flexibility was designed into the plant and the eventual process configuration to be utilized will be selected on an empirical basis. There is some apprehension regarding the energy costs for aerobic digestion which was designed into the plant as an option prior to the surge in energy prices. If aerobic digestion proves too costly, anaerobic digestion of mixed sludges will be evaluated.

## DESIGN EXAMPLE—DRYING BEDS—60 MGAL/D (2.63 $m^3/s$) PLANT

The design of the new Tampa AWT plants' drying beds serves as an example of the design of this type system for a large plant in a subtropical climate.

Estimates of quantities of unstabilized sludge solids to be encountered in the new plant are summarized in table 6–4.

Faced with processing the daily volumes of sludges shown and considering the acceptable results previously achieved at Tampa with anaerobic digestion, further calculations of the amounts of sludges which would result from anaerobic digestion of primary solids and aerobic digestion of AWT solids were carried out and results are listed in table 6–5.

Table 6–4.—Tampa AWT plant, estimated annual average unstabilized byproduct solids production

| Item | Year | |
|---|---|---|
| | 1976 | 1985 |
| Primary solids slurry | | |
| lbs/day (dry) | 37,000 | 37,000 |
| percent solids | 5.0 | 5.0 |
| gals/day | 89,000 | 89,000 |
| AWT solids slurry | | |
| Biological solids | | |
| lbs/day (dry) | 44,000 | 71,000 |
| Chemical solids | | |
| lbs/day (dry) | 31,000 | 48,000 |
| Total | 75,000 | 119,000 |
| percent solids | 3.0 | 3.0 |
| gals/day | 300,000 | 476,000 |
| Combined solids slurry | | |
| lbs/day (dry) | 112,000 | 156,000 |
| percent solids | 3.5 | 3.3 |
| gals/day | 389,000 | 565,000 |

Table 6–5.—Tampa AWT plant, estimated annual average stabilized byproduct solids production

| Item | Year | |
|---|---|---|
| | 1976 | 1985 |
| Primary solids slurry | | |
| lbs/day (dry) | 14,000 | 14,000 |
| percent solids | 3.0 | 3.0 |
| gals/day | 56,000 | 56,000 |
| AWT solids slurry | | |
| Biological solids | | |
| lbs/day (dry) | 38,500 | 57,500 |
| Chemical solids | | |
| lbs/day (dry) | 31,000 | 48,000 |
| Total | 69,500 | 105,500 |
| percent solids | 5.0 | 5.0 |
| gals/day | 169,000 | 253,000 |
| Combined solids slurry | | |
| lbs/day (dry) | 83,500 | 119,500 |
| percent solids | 4.4 | 4.6 |
| gals/day | 225,000 | 309,000 |

A series of sludge solids stabilization, dewatering, and disposal options were then reviewed for reliability, environmental impact, and capital plus operating and maintenance costs. Table 6–6 summarizes these cost results.

Based on the comparative costs shown and on other

Table 6–6.—Tampa AWT plant, alternative byproduct solids systems total cost comparison

| Rank | Description | Estimated comparative costs—$1,000,000 Capital | Average annual | Average annual cost per ton raw solids[a] |
|---|---|---|---|---|
| 1 | Air dry—with chemicals—cake to user | $11.67 | $2.75 | $96.52 |
| 2 | Air dry—cake to user | 14.14 | 2.84 | 99.81 |
| 3 | Air dry—with chemicals—cake to landfill | 11.67 | 3.31 | 116.16 |
| 4 | Air dry—cake to landfill | 14.14 | 3.40 | 119.46 |
| 5 | Kiln dry—without anaerobic digestion | 15.18 | [b]3.44 | [b]120.68 |
| 6 | Kiln dry—with anaerobic digestion | 16.07 | [b]3.50 | [b]122.76 |
| 7 | Mechanical dewatering | 15.87 | 3.84 | 134.85 |
| 8 | Liquid spray | 23.79 | 4.32 | 151.78 |
| 9 | Liquid slurry to user | 23.65 | 4.38 | 153.79 |
| 10 | Incineration | 21.47 | 4.71 | 165.49 |

[a]Based on 78 tons per day (dry) raw byproduct solids.
[b]Net after revenue deduction from sale of product.

Table 6–7.—Tampa AWT plant, design criteria—drying beds

| Air drying beds | Design year—1985 Annual average | Maximum month |
|---|---|---|
| Volume each drying bed (gals at 12″ fill depth) | 65,000 | 65,000 |
| Area each drying bed (ft$^2$) | 8,690 | 8,690 |
| Number of drying beds | 140 | 140 |
| Total area (ft$^2$) | 1,216,600 | 1,216,600 |
| Drying time (days) | 29.5 | 19.6 |
| Solids loading (lbs/ft$^2$/yr) | 35.85 | 53.79 |
| Dried solids | | |
| lbs/day (dry) | 119,500 | 179,300 |
| percent solids | 40.0 | 40.0 |
| lbs/day (wet) | 298,800 | 448,300 |
| tons/day (wet) | 149 | 224 |
| cu ft/day (wet) | 3,900 | 5,800 |

evaluation factors, the alternate of air drying (drying beds) with use of flocculants was chosen as the most cost effective.

The total estimated capital cost for the air drying system being installed at Tampa currently, including all piping, auxiliaries such as equalizing storage, site work, engineering, underdrainage system, etc., was $4,671,000 including $941,000 contingency.

The drying bed operational design criteria are as shown in table 6–7.

## PAVED RECTANGULAR DRYING BEDS WITH CENTER DRAINAGE

A good example of this type of system is that at Dunedin, Fla. Figure 6–3 is a photograph of the Dunedin beds.

As can be seen, the two beds in the left portion of the photo contain previously loaded sludge which is drying. The two empty beds on the right are ready to be loaded.

The Dunedin plant is of interest due to use of a unique heated drying bed system.

Plant process features:

1. An average flow of 2.5 Mgal/d (.11 m$^3$/s) of primarily domestic wastes.
2. Liquid treatment via primary sedimentation followed by conventional activated sludge. The plant originally used a contact stabilization system but was converted to conventional activated sludge with positive results.
3. Primary sludge is subjected to two stage anaerobic digestion with a Pearth gas recirculation system.
4. The excess activated sludge is thickened in a DAF unit and most of the thickened WAS then goes into the anaerobic digester system. Some of the WAS is subjected to aerobic digestion, but no more than necessary due to the energy consumption of same.

Figure 6–3.—Paved rectangular heated drying beds, Dunedin, Fla.

(The operation of the DAF unit is well managed, as is the entire plant, and the plant is a good reference point for the proper application of DAF thickening in a smaller plant.)

5. The digested sludges are processed in three different ways. A portion is dried on the heated drying beds prior to use as a soil additive. Some of the sludges are dewatered on an existing rotary vacuum filter when this is required. An additional portion is disposed of in liquid form via tanker.
6. The digester gas is burned in a hot water heating system. The heated water is circulated through piping in the paved portion of the drying beds.

The Dunedin plant has four drying beds (75×25 feet each) (22.86×7.62 m) or 7500 ft$^2$ (696.8 m$^2$) of *evaporative* drying area. The *drainage* drying area, due to the type of construction is only a fraction of the evaporative area. The beds are heated, as noted, but are not covered and the Tampa Bay area has a high average annual rainfall. Polyelectrolytes are used to condition the sludge.

Sludge drying time (averages) to liftable condition is 5 days normally and 12 days in rainy periods. The beds are charged with 5,000 gallons (13,930 l) of a 2.6 percent dry solids content sludge at a time. Thus the loading rate varies from 18 to 43 pounds (6.35 to 19.50 kg) of dry solids sludge per square foot (.09 m$^2$) per year.

With a 5-day drying period the 4 beds are capable of dewatering about 13 dry tons (11.8 Mg) per month. Certainly the capacity of 43 pounds per square foot per annum (209.8 kg/m$^2$/year) achieved at Dunedin is several times greater than the Ten States Standards for conventional open beds.

## WEDGEWATER DRYING BEDS

Wedgewater "Filter Beds" or drying beds were designed to introduce sludge slurry onto a horizontal relatively open drainage media in a fashion which would yield a clean filtrate and also give a reasonable drainage rate.

The Wedgewater Filter Bed (figure 6–4) consists of a shallow rectangular watertight basin fitted with a false floor of wedgewater panels. These panels have slotted openings of 1/4 MM and produce a total open area of 8 percent. The boundary of this false floor is made watertight with caulking where the panels abut the walls. An outlet valve is fitted in one wall of the bed to communicate with the underside of the wedgewater decking.

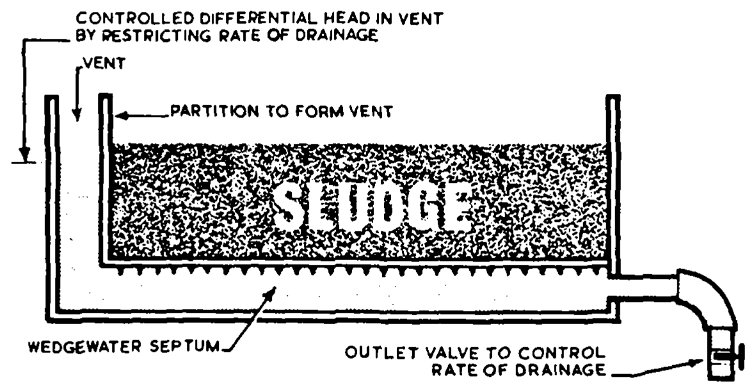


Figure 6–4.—Cross section of a wedgewater drying bed.

The controlled drainage rate is obtained by first introducing a layer of water into the wedgewater unit to a level above the septum. The sludge is then slowly introduced and in effect, under the proper conditions, floats on the water layer. After the proper amount of sludge has been introduced, the initial separate water layer and drainage water is allowed to percolate away at a controlled rate. The exact procedure varies somewhat with different types of sludges. It is apparent that for this concept to perform as intended the sludge and the initial water layer must be relatively immiscible.

The wedgewater technique is designed to permit controlled formation of a cake at the crucial sludge/support media interface before any significant quantity of fines migrates to the interface or into the openings of the septum or escapes in the filtrate. Since polyelectrolyte flocculants promote rapid cake formation and bind up fines they are now used in conjunction with Wedgewater Filter Bed installations processing municipal sludges.

Each square foot (0.9 m) of wedgewater can normally dewater between 1/2 lb (.23 kg) and 1 lb (.45 kg) of dry matter per charge. The loading rate depends on the initial solids concentration of the waste sludge applied. Most sludges can be dewatered to a handleable condition of 8 to 12 percent solids within 24 hours. This process is most practical for the smaller treatment plant which has an average daily flow of 500,000 gal/day (.02 m$^3$/s) or less. Sludge loading rates of 182–365 lb/ft$^2$ per year (882.2–1781.2 kg/m$^2$/year) are normal.

Results with Wedgewater units at 2 U.S. plants are described in the following paragraphs.

## ROLLINSFORD, N.H.

This plant produces an excess biological sludge at the rate of 150 gallons per day (567 l) at 2 percent dry solids content. A wedgewater unit, as shown in figure 6–5, is used to dewater the sludge to a solids content of 8 percent, which is liftable.

A polyelectrolyte conditioner is used in the process. Calculations from the data in the reference cited show that conservatively assuming 2 drying cycles per day for the 15′ by 6′ (4.57×1.83 m) unit, the production rate could be 1.1 lb/hr/sq ft (5.4 kg/m$^2$/hr), or 570 lb/sq ft year (2780 kg/m$^2$/year) which is, of course, an order of magnitude greater than the dewatering rates normally associated with conventional drying beds. These results are tempered by the fact that 8 percent, while a liftable condition for this sludge, is not a particularly high solids content. It is apparent, however, that higher than 8 percent solids would be readily obtainable with increased drying times while still maintaining a very high annual solids loading, if such a higher solids content were required.

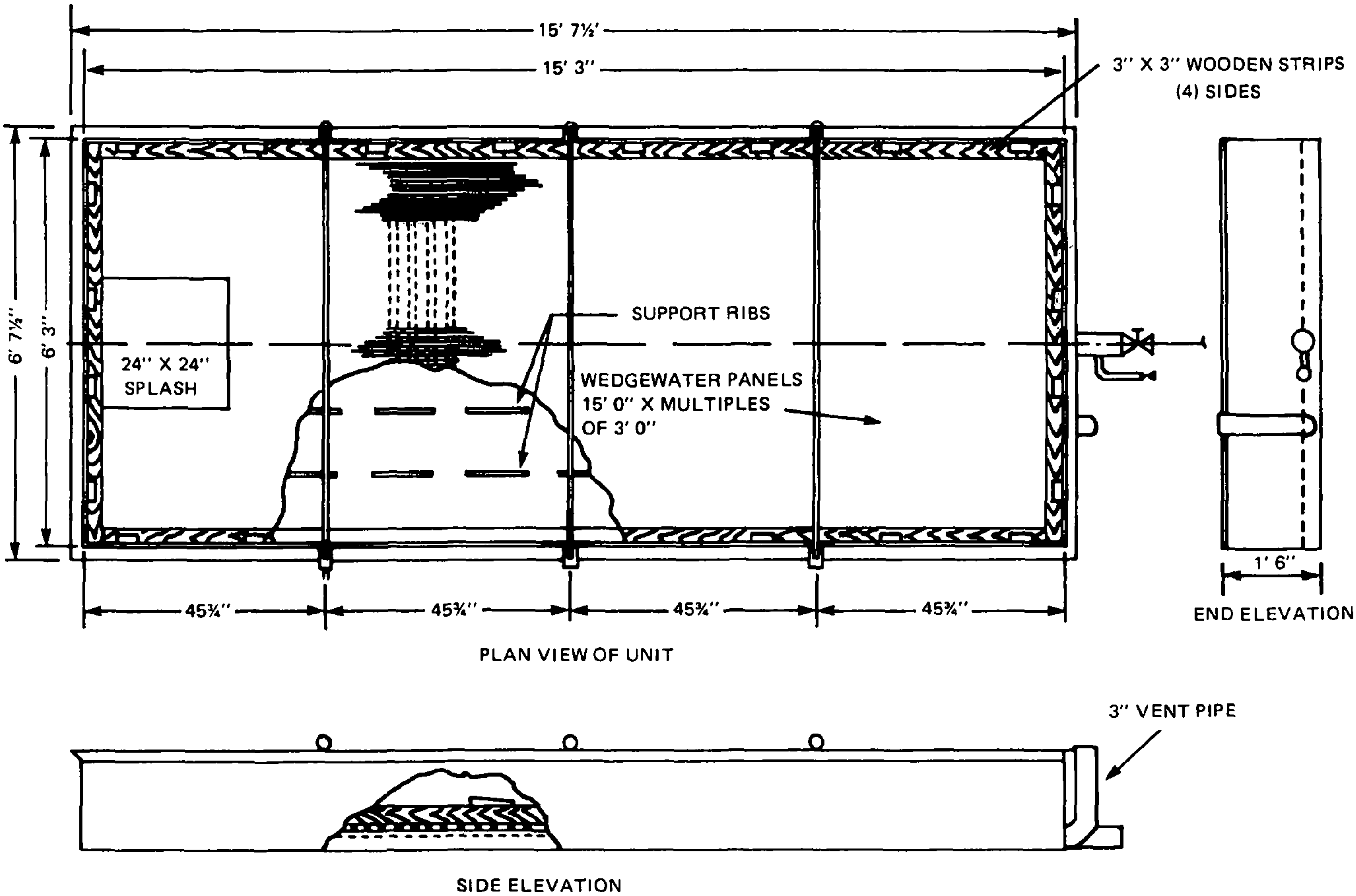


Figure 6–5.—Wedgewater drying bed, Rollinsford, N.H.

## DUNEDIN, FLA.

Additional results on the wedgewater system are reported from work at Dunedin, Fla. At that location, the biological sludge was dewatered to a solids content of 10.4 percent in 22 hours through the mechanism of the wedgewater element, use of support water, and the restricted drainage procedure, without the use of polymer flocculants.

There are 18 U.S. installations of the wedgewater system. Several are industrial applications but most are installed at small plants of the contact stabilization type.

A tiltable unit, more or less similar to the lift and dump mechanism of a dump truck, is available to facilitate removal of sludge when slightly fluid cake can be tolerated or when removal by rake is feasible. The supplier, Hendricks Manufacturing Co. of Carbondale, Pa., also supplies design recommendations for mechanical removal via small front end loader when indicated. A 1 square foot (0.09 $m^2$) bench scale test model is available for test purposes.

The stainless steel wedgewire septum in the 15- by 6-foot (4.57 × 1.83 m) Rollinsford unit would cost $4,500 at today's prices.

## VACUUM ASSISTED DRYING BEDS

At the 4.5 Mgal/d (.20 $m^3$/s) Sunrise City, Fla., contact stabilization plant, a vacuum assisted drying bed system has been used for the past 18 months to dewater the 2 percent dry solids sludge. Principal components of the system are:

1. A rigid multimedia filter top surface.
2. An intermediate void filled with stabilized aggregate.
3. A low impermeable barrier, consisting of reinforced concrete. (It would alternatively be pre-fabricated fiberglass.)

Figure 6–6 is a photograph of one of the two drying bed units showing the sludge being fed onto the surface of the upper multimedia in one of the beds.

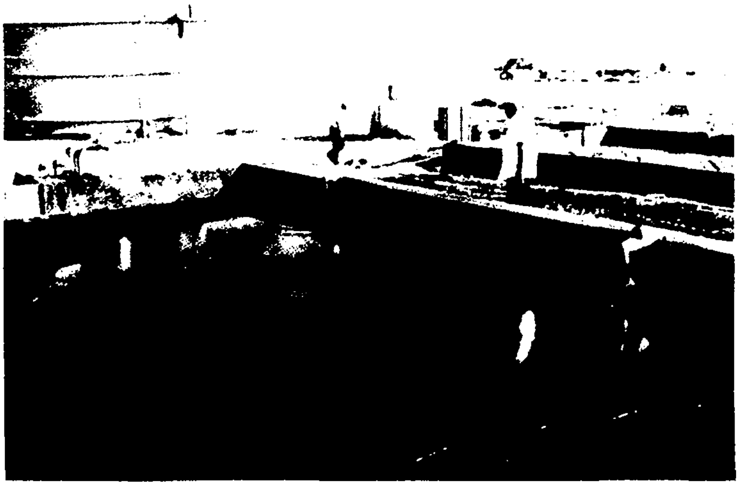

Figure 6–6.—Rapid sludge dewatering beds, Sunrise City, Fla.

The following sequence of operations is used:

1. Sludge is fed onto the filter surface by gravity flow at a rate of 150 gallons (567 l) per minute to a depth of 12 to 18 inches (.30 to .46 m).
2. Filtrate is drained through the interconnected voids of the stabilized aggregate to a sump, from which it is pumped back to the plant by a self-actuated submersible pump.
3. As soon as the entire surface of the rigid filter is covered with sludge, the vacuum system is turned on to maintain a vacuum of one to 10 inches (2.54 to 25.4 cm) of mercury on the intermediate void area.

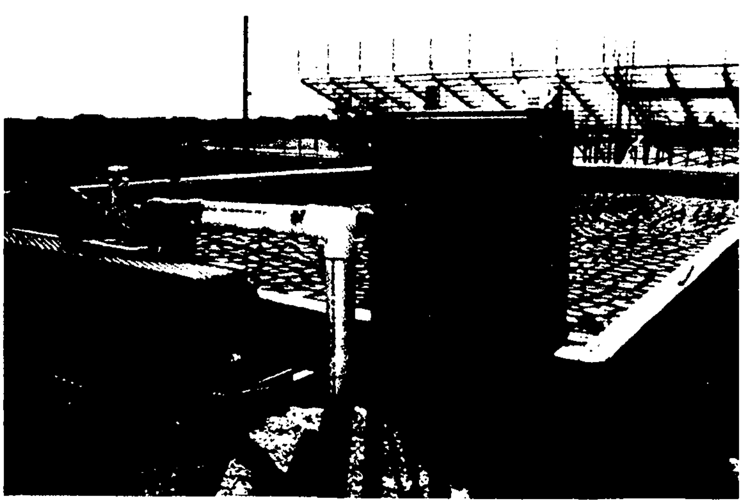

Figure 6–7.—Vacuum assisted drying beds, Sunrise City, Fla.

Under favorable weather conditions, this system dewaters the 2 percent solids aerobically digested contact stabilization sludge (a difficult high bound water content sludge) to a 12 percent solids level in 24 hours without polymer use, and to the same level in 8 hours if flocculant is used. The 12 percent condition is liftable. The sludge will further dewater to about 20 percent solids in 48 hours.

The sludge cake is removed from the filter surface either manually, mechanically by a small hydrostatic drive front-end loader such as a Melroe Bobcat 520, or by a vacuum truck.

Controlled tests of this type system have shown that a sludge loading rate of 306 pounds per square foot year (1490 kg/m$^2$/year) is attainable.

At Sunrise City plant (figure 6–7), the two 20 feet by 40 feet (6.10 × 12.19 m) vacuum drying beds are processing a substantial portion of the total plant load. The photograph below shows the appearance of a bed at the end of the drying period and also shows the proximity to a local athletic field.

The vacuum assisted drying bed system at Sunrise City is a proprietary system now designed and supplied by International Sludge Reduction Co.

## DESIGN EXAMPLE—DRYING BED FOR 4 MGAL/D (.18 m$^3$/s) PLANT

### Basic Assumptions

These assumptions are as follows:

1. The sludge to be processed is an anaerobically digested mixture of primary and WAS at 4 percent dry solids content. It is a mixture of 60 percent primary sludge and 40 percent WAS with the WAS originating from a conventional activated sludge system.
2. Ultimate disposal is to be by hauling to a sanitary landfill, or to farmland or other horticultural use.
3. Equilibrium sludge removal rate of 2.5 tons (2.3 Mg) of dry solids per day to be maintained.
4. The plant is located in the Middle-Atlantic section of the United States.

### Alternate Units for Consideration or Evaluation

For a plant of this size, depending on site limitations, either conventional enclosed drying beds or vacuum assisted enclosed drying beds should be considered. The economics and other constraints of final disposal, such as length of truck haul and final solids content requirements would bear consideration. Land area availability would materially affect the choice between gravity or vacuum assisted drying beds. If excess methane was available from anaerobic digestion, consideration could be given to use for heating the enclosed bed air space during the winter.

For the purposes of this example it is assumed that sufficient land area is available for either gravity or vacuum assisted drying beds.

## Evaluation Procedure

The general sequential procedure recommended to be followed would be similar to that fully described on page 37 in the RVF design example. The only variation would be that bench scale and/or pilot plant tests on the drying bed dewatering characteristics of the sludge would probably have to be planned and carried out entirely by the consulting engineering firm and the client for the conventional enclosed bed option. On the vacuum assisted bed option the suppliers have developed small scale testing procedures and could be involved in the work.

## Pilot Scale Tests

Since temperature conditions could affect the sizing of enclosed beds it is suggested that, in the absence of available data from existing plants in the same general area with equivalent sludges, a small greenhouse type test installation would be in order. Ready-made unitized small greenhouse enclosures intended for the homeowner are now available at modest prices and could be adapted to enclose a small drying bed for test work on both options.

## Design Calculations

It is assumed that the test work has shown that by enclosing the beds and using in line flocculant conditioning the average bed loading for the conventional gravity system is 55 lb/ft$^2$ per year (268 kg/m$^2$/year) and for the vacuum assisted option is 110 lb/ft$^2$ per year (537 kg/m$^2$/year).

1. Since drying bed operation is a batchwise procedure a sludge storage or surge vessel should be provided to contain the thickened digested sludge and serve as a feed tank for the drying beds.
2. Sludge volume rate would be 14,000 gallons/day (53,000 l/d) or 98,000 gallons (371,000 l) per week, so a 100,000 gallon (378,500 l) surge vessel would be required as a feed tank.
3. Assuming tests showed a 12 inch (0.30 m) bed fill level to be practical, for the conventional gravity beds loaded at a conservative loading of 47 lb/ft$^2$ per year (229 kg/m$^2$/year), five beds, each 65 feet by 120 feet (19.81 × 36.58 m) would be adequate.
4. The use of five beds would permit the bed filling procedure to average less than two per week on an annual basis.
5. For the vacuum assisted bed option using a conservative design loading of 91 lb/ft$^2$ per year (444 kg/m$^2$/year) would result in selection of four 50 feet by 100 feet (15.24 by 30.48 m) drying beds.

## Additional Considerations

The system should include for either of the two options, mechanical sludge removal via a front end loader.

An important point in evaluating the two options would be a determination of the energy requirements involved in operating the vacuum system in that option.

# FUTURE OF DRYING BEDS

An objective review of past results and consideration of the developments of the past 5 to 7 years in modifying and increasing the dewatering capacity and improving the mechanical removal capabilities of drying beds must lead to the conclusion that they should be much more widely used than at present.

It seems clear that a judicious combination of the following aspects would in many locations make drying beds the dewatering system of choice:

1. Provision in the bed design for mechanical removal via front end loaders a la Tampa, etc.
2. Provision for conditioning of the sludge on its way into the bed with polyelectrolytes or equivalent as needed.
3. Inclusion in the design of a translucent roof, or a total greenhouse type enclosure with adequate ventilation and odor control systems.
4. Where required for capacity purposes some form of vacuum assistance (a la Sunrise City, Fla.) for increasing the drainage rate and enhancing evaporation where indicated.

If these aspects were included in conceptual designs, the design criteria in terms of square footage of bed area required would be many times less than the figures listed in the Ten State Standards. As a result of this an overall system evaluation of cost-effectiveness would surely result in more widespread use of drying beds than is currently the case.

# ROTARY VACUUM FILTERS

There are three normal types of rotary vacuum filters and they are described in table 6–8.

The first (drum) type was largely displaced by the latter two due to cloth plugging problems associated

Table 6–8.—Types of rotary vacuum filters

| Type | Support media | Discharge mechanism |
|---|---|---|
| Drum | Cloth | Blowback section/doctor blade |
| Coil | Stainless steel coils | Coil layer separation/tines |
| Belt | Cloth | Small diameter roll, flappers, doctor blades |

with the use of lime and ferric chloride/lime conditioning systems. The drum type filter does not exhibit cloth plugging problems with polyelectrolyte flocculants.

The coil filter has been widely used and does have a positive release mechanism. Care must be exercised with coil filters to ensure a sufficiently rapid rate of cake formation to prevent loss of fines through the more open media involved during the initial phase of cake formation. This is a relatively infrequent problem and if the fines problem does occur it is usually symptomatic of predewatering processes which have destroyed a substantial portion of the matrix forming material in the sludge(s) or of inadequate conditioning. Such pretreatment processes will be detrimental in some manner to any dewatering device.

Belt type filters were introduced to permit continuous washing of the cloth and ostensibly overcome effects of plugging by lime or fines. This concept was erroneous in most cases since the belt washes were not particularly effective in removing lime. In several plants which had early installations of the Drum type filter and later installations of Belt filters side by side, the purported advantages of the Belt filters proved to be illusory. Belt type filters are particularly prone to cake discharge problems.

Rotary vacuum filters produce typical results when inorganic chemicals are used for conditioning. The results appear in table 6–9.

While the data in this table above and the following one are representative, they should not be used for design purposes if the actual sludges to be dewatered are available for lab and/or pilot test work. It should also be noted that the cake solids figures shown in this table include the significant amounts of ferric chloride and lime used so the actual sewage solids content is lower than what is shown. For instance, the correction would typically bring the net sewage solids of a 22 percent cake down to a correct figure of 18 percent.

There are instances where a combination of ferric chloride and polyelectrolyte is employed to maximize rotary vacuum filter production rate. This is frequently the case where the sludge has a high grease content and tends to stick to the filter cloth on belt type filters.

Table 6–9.—Typical rotary vacuum filter results for sludge conditioned with inorganic chemicals

| Type sludge | Chemical dose (percent) | | Yield (lb/hr/ft$^2$) | Cake solids percent |
|---|---|---|---|---|
| | Ferric chloride | Lime | | |
| Raw primary | 1–2 | 6–8 | 6–8 | 25–38 |
| Anaerobically digested primary | 1–3 | 6–10 | 5–8 | 25–32 |
| Primary and humus | 1–2 | 6–8 | 4–6 | 20–30 |
| Primary and air activated | 2–4 | 7–10 | 4–5 | 16–25 |
| Primary and oxygen activated | 2–3 | 6–8 | 5–6 | 20–28 |
| Digested primary and air activated | 4–6 | 6–19 | 4–5 | 14–22 |

Table 6–10.—Typical rotary vacuum filter results for polyelectrolyte conditioned sludges

| Type sludge | Chemical cost ($/ton) | Yield (lb/hr/ft$^2$) | Cake solids |
|---|---|---|---|
| Raw primary | 1.5–3 | 8–10 | 25–38 |
| Anaerobically digested primary | 3–6 | 7–8 | 25–32 |
| Primary and humus | 4–8 | 4–6 | 20–30 |
| Primary and air activated | 5–18 | 4–5 | 16–25 |
| Primary and oxygen activated | 5–15 | 4–6 | 20–28 |
| Anaerobically digested primary and air activated | 6–22 | 3.5–6 | 14–22 |

Aluminum chloride or aluminum chlorohydrate are also effective inorganic conditioning agents and where plants have existing rotary vacuum filters, the availability of such materials as waste byproducts of industrial plants is worth exploration.

Typical results for polyelectrolyte conditioned sludges are described in table 6–10.

In point of fact, more of the sludge processed in plants equipped with rotary vacuum filters is conditioned with polymer flocculants than with inorganic conditioners. The chemical cost is normally about the same for the use of polyelectrolytes or inorganic conditioners. The use of polyelectrolytes largely prevails because of more convenient handling, less extensive preparation facilities, and freedom from corrosion problems, plus the elimination of significant quantities of inorganic solids in the dewatered cake.

On the other hand, some plants must use inorganic conditioners to obtain cake release, provide matrix forming material in the cake, or to facilitate lime addition for ultimate disposal.

With a digested mixture of primary and excess activated sludge, in most plants, rotary vacuum filters will produce dewatered cakes with cake solids contents within the 18–22 percent range, which is almost always too wet for autogenous incineration or some composting processes. These facts, plus energy costs have caused the selection rate for rotary vacuum filters to wane considerably.

The sludge feed to rotary vacuum filters should never be below 3 percent dry solids content and preferably should be greater than 4 percent if reasonable production rates are to be attained.

## AUXILIARY DEVICES FOR ROTARY VACUUM FILTERS

To obtain higher solids cakes from rotary vacuum filters (RVF), three companies have developed devices which can further dewater the filter cake. These devices are, in some cases, specifically designed as add-ons to existing filters or in others, supplied as integral parts of the rotary vacuum filter.

The items of reference are:

1. The Eimco Hi-Solids filter.
2. The Parkson Magnum Press high pressure section.
3. The Komline Sanderson Unimat high pressure section.

## Eimco Hi-Solids Filter

This device combines normal rotary vacuum filtration with a batch type adjunct pressure filter. The cake while still on the rotary vacuum filter belt feeds into a small co-joined stage where it is subjected on one side to pressure from a rubber diaphragm (50–150 lb/in.$^2$ g or 3.5–10.5 Kg/cm$^2$) while on the other side (below the belt) a vacuum is applied to facilitate drainage. Since this is a batch procedure with the rotation of the rotary vacuum filter being momentarily interrupted while the pressure and vacuum are applied in the pressure chamber section, some lowering of production occurs.

Eimco supplies this unit as an integral system and also supplies the press portion as an add-on device for existing conventional rotary vacuum filters. This device was tested on pilot scale at Washington, D.C., and increased the cake solids content from a normal 17 percent up to a level of 25 percent. The sludge tested was a rather difficult to process mixture of primary and secondary sludges.

## Parkson Magnum Press

This unit (more fully described in the section on Continuous Horizontal Belt Filters) was evaluated on pilot scale at Washington, D.C., for dewatering filter cake from the existing rotary vacuum filters. Filter cake of 18 percent dry solids content was further dewatered to 35–40 percent dry solids with no further conditioning employed.

Commercial availability of this unit hinges on successful conclusion of development work required to enable design of a mechanical method of transmitting filter cake from the rotary vacuum filter to the auxiliary press section without degrading the processability of the cake.

## Komline Sanderson Unimat

A pilot model of the medium and high pressure sections of the Unimat was evaluated at Washington, D.C., on the cake from the rotary vacuum filters and produced a cake of 38 percent dry solids. Once again mechanical development work would be required to facilitate an installation.

In summation, the three devices briefly described above offer real promise for providing a means to further dewater sludge cake from existing rotary vacuum filter installations where such a procedure is in order.

# DESIGN EXAMPLE—ROTARY VACUUM FILTRATION 4 MGAL/D (0.18 m$^3$/s) PLANT

## Basic System Assumptions:

The sludge is an anaerobically digested mixture of primary and excess activated sludge which has been thickened to 4 percent solids via a flocculant aided post-digestion thickening process. System design has been such that the sludge mixture is about 60 percent primary and 40 percent secondary sludge. The sludge is available for testing.

The ultimate disposal method for the sludge is to be by hauling dewatered cake to either a sanitary landfill, or for disposal on farmland, or for composting and horticultural use.

The sludge removal rate required is to average 2.5 dry tons (2.3 Mg) per day and the cake must possess sufficient dimensional stability to preclude flow out of a truck.

## Alternate Units for Consideration and/or Evaluation

1. A Coil filter.
2. A Belt type filter.
3. A Drum type filter.

## Evaluation Procedure

The sequence to be followed in the evaluation and design is planned as follows:

1. Verification of the amounts and relative degree of uniformity of the flow of sludge to be dewatered. This is to be obtained by review of plant operating data.
2. Diagnostic bench scale dewatering tests of the sludge, repeated several times during different operational periods to assure uniformity. It is absolutely essential that these tests and any pilot tests be done on site with fresh sludge.
3. Review of the above results with interested candidate suppliers and then repetition of the bench scale tests in conjunction with suppliers personnel.
4. A pilot dewatering test series should then ensue, particularly if there is any doubt about any facet of the dewatering operation. This should be carried out with at least two of the potential suppliers.
5. Summation of design data should be prepared by the consulting engineer. Each potential supplier should be asked to prepare and transmit a report of the bench and pilot test work including their design recommendations, including equipment required, sizing, delivery time, etc., together with "budget price quotes" and estimates of annual operation maintenance costs, and life cycles of the various items of equipment.
6. A detailed design should then be prepared and plans, specifications, conditions of contract, etc.,

forwarded to those suppliers whose equipment and performance have qualified them to enter a firm price quotation.

7. From the design and overall system cost data available, and with full consideration of relative equipment reliabilities, a selection of the supplier can then be made.

## Bench Scale Tests

The "Buechner Funnel" test procedure is well documented and all suppliers of rotary vacuum filters are very familiar with it. The "Filter Leaf" test procedure is likewise readily available.

Normally the Buechner Funnel test, employing a cake support media identical to that to be employed will supply all the required information needed. However, if the dewatered cake shows real signs of sticking to the filter media, then a leaf test to check this property may be in order.

In the Buechner Funnel test it is important to:

1. Determine dewatering rate, time to vacuum break and resultant cake solids after a simulated cycle.
2. Analyze the filtrate for suspended solids, $BOD_5$, COD, and total dissolved solids.
3. The data from (2), along with analogous sludge feed data should be used to determine exactly what total solids capture is being obtained.
4. The cake release characteristics should be carefully assessed. If a problem is indicated, a leaf test can be run to observe whether or not the cake falls freely from a vertically held leaf. If it doesn't, then a Belt filter will cause release problems.

## Pilot Tests

Most suppliers have packaged pilot units which can be wheeled in for testing. This is advisable, in most cases.

It is important that the sludge quality during the comparative pilot plant tests be reasonably comparable. This can be verified by concurrent "Buechner Funnel" testing.

## Design Calculations

1. Operating cycle to be 35 hours per week (7 hours/day). This permits start-up and wash-up times within an 8 hour shift
2. One filter, with adequate supply of key spare parts to be maintained.
3. Size of vacuum filter.—Production rate has been determined via pilot testing to be 5 lb/hr/ft$^2$ (24 kg/m$^2$/hr), but to provide a margin of safety, 4 lb/hr/ft$^2$ (20 kg/m$^2$/hr), will be used. Steady state sludge removal rate requirement is 35,000 pounds (15,870 kg) per week. With a 35 hour-per-week schedule, weekly filter capacity at 4 pounds per hour per square foot (20 kg/m$^2$/hr) is 140 pounds (63.5 kg) per ft$^2$ 35,000 pounds/week (15,870 kg/wk) ÷ 140 pounds (lbs/ft$^2$/wk) per square foot per week (685 kg/m$^2$/wk) = 279 square feet (26 m$^2$) of filter area required. The nearest standard size filter is 300 square feet (28 m$^2$), so a single unit of this size is chosen.
4. Sizing of auxiliary equipment.—In each case the details of sizes of vacuum equipment, conveyors or other system required to get the dewatered cake into the truck for hauling, and the chemical dosing equipment for sludge conditioning must be developed, and priced.
5. Sludge storage capability.—The one shift per day—five day per week mode of operation plus the use of a single filter will require provision of several days storage capacity for the digested sludge. This could potentially be provided by a combination of the inherent surge capacities of the digestion tanks and post digestion thickening tanks, or by provision of a separate storage tank equipped to ensure homogeneity of feed to the RVF.

For a sludge of the type described, a cationic polyelectrolyte flocculant would probably be used for conditioning. The testing and selection of suitable conditioning agents would necessarily be carried out in conjunction with the series of bench scale and pilot test programs used to select and size the rotary vacuum filters. As part of the selection process for suitable conditioners, data should be obtained and reviewed on:

1. Price, dosage rate, and availability of both polyelectrolytes and inorganic conditioners in the particular locale.
2. The system required for solution preparation and application, and its cost.
3. The storage stability (shelf life) of the conditioner in its form as supplied and in stock solution for use.
4. Handling characteristics, safety aspects and corrosion properties of the material in dry and liquid forms.
5. Previous experience with the same materials at other plants with similar sludges.

# DEWATERING SYSTEM CONSIDERATIONS

Auxiliary equipment such as sludge conveyor or removal facilities, chemical mixing and feed equipment, and sludge feed pumps are usually available from the rotary vacuum filter supplier.

Polymer solution preparation and dosing equipment is also frequently available from the polymer supplier or from an equipment supplier other than the rotary vacuum filter supplier.

An Energy Audit should be a part of every system evaluation. The Energy Audit should include not only an estimate of the power consumption of the dewatering equipment and its immediate auxiliaries, but also the impact of the particular dewatering system on the overall treatment process system. In this regard, the assessment should specifically include the impact of the conditioning/dewatering system on both the post dewatering portion of the system and the pre-dewatering portion of the system. The latter facet makes preparation and consideration of "Quantified Flow Diagrams" for both the conditioning/dewatering system and the overall treatment system mandatory to cost effective design.

For purposes of comparison, the rotary vacuum filter in this design example would require a vacuum pump of 30 horsepower (22 kW), and filtrate pump of 3 horsepower (2.2 kW). To make a complete energy audit, all the auxiliary equipment data, and the other points mentioned in the previous paragraph would have to be assessed.

## DESIGN EXAMPLE—ROTARY VACUUM FILTRATION—40 MGAL/D (1.75 m³/s) PLANT

### Basic System Assumptions

These would be the same as in the preceding design example for a 4 Mgal/d (0.18 $m^3/s$) plant except that the required removal rate would be 25 tons of dry solids per day (22.7 Mg/day).

### Other Considerations

The following parts of the Design Example would be the same as for the 4 Mgal/d plant (0.18 $m^3/s$) in the preceding example:

1. Alternate units for consideration and evaluation.
2. Evaluation procedures.
3. Bench scale testing.
4. Pilot tests.

### Design Calculations

1. Operating cycle.—To be either a seven day per week, 24 hour per day operation or five day per week, 24 hour per day operation depending on reduction and final disposal processes chosen.
2. Size and number of Rotary Vacuum Filters required.—Production rate to be conservatively taken at 4 pounds/hour/sq ft (20 kg/hr/$m^2$). At 350,000 pounds (159 Mg) per week the weekly capacity of a square foot of filter area for a seven day operation (allowing 2 hours/day downtime average for clean up and maintenance) is 4 pounds/hour/sq ft (20 kg/hr/$m^2$)$\times$154 hours per week or 616 pounds/week/sq ft Dividing 350,000 pounds per week by 616 (3) gives a filtration area requirement of 568 square feet (53 $m^2$). A similar calculation for a five day operation gives a filtration area requirement of 793 square feet (74 $m^2$). In either the seven day/week or five day/week options, two 500 square foot (46 $m^2$) rotary vacuum filters would normally be specified to provide sufficient capacity and redundancy.
3. All of the other facets of the design procedure would be the same as in the 4 Mgal/d (0.18 $m^3/s$) example.

### General Comment—Rotary Vacuum Filters

The RVF was, for many years, the common device for dewatering municipal sludges. Their frequency of use had persisted longer in the United States than in the rest of the world.

Operating problems such as the cake pick-up difficulties, poor cake release from belt filters with sticky sludges, and the maintenance requirements associated with vacuum producing equipment have existed in numerous cases. Solids capture problems associated with either the effect of less than adequate cake formation rate in some relatively open media filter installations or with cake recycle due to sticking problems have also occurred. While these problems could be moderated in many cases by revision of conditioning methodology or mechanical changes, they are deterrents to widespread continued usage.

More universal deterrents to the continued selection of RVF's are:

1. The energy and maintenance costs associated with operating vacuum systems.
2. The inability to produce nearly as dry a cake as other newer devices.

These comments are made to encourage the design engineer to review current operating and cost experiences at existing plants prior to making a design decision.

## CONTINUOUS BELT FILTER PRESSES

This general type of device, which employs single and/or double moving belts to continuously dewater sludges through one or more phases of dewatering was originally developed, and in subsquent years modified and improved, in West Germany. The earliest concurrent U.S. development was under the aegis of the late Brian Goodman, at Smith and Loveless Division of Ecodyne.

The scope and depth of development of this newer type device has been much more pronounced in Europe than in the United States until the past 3 to 4 years. Within those past 3 to 4 years, many different models of the same type device, differing in configuration and capability, have been introduced into the U.S. market.

While there is general agreement that the Continuous Belt Filter Press (CBFP) materially extends capabilities for improved dewatering of sludges, the U.S. design engineer is faced with a real task in selecting the optimum device from the many which are now available. But that task must be dealt with if advantage is to be taken of this technological breakthrough.

U.S. installations of the latest and best models are just now coming onstream. To review actual operating performance on particular sludges, usage of available mobile pilot test units, coupled with site visits is in order. There is considerable operating experience available at existing European sites. The old conundrum that European sludges are different and results are not applicable should be treated with the contempt it deserves, since it is inaccurate.

## Original Concept and Evolutionary Developments—Continuous Belt Filter Presses

Figure 6–8 illustrates the single level device originally marketed by Klein of Germany and their U.S. licensee, R. B. Carter.

Practically concurrent with this development was Brian Goodman's work with the Smith & Loveless Concentrator which is described later.

This type unit was successful with many normal mixed sludges. Typical dewatering results for digested mixed sludges with initial feed solids of 5 percent are to give a dewatered cake of 19 percent solids at a rate of 6.7 lb/hr/sq ft (32.8 kg/hr/m$^2$) with a chemical conditioning cost of \$4.10/ton (\$4.52/Mg). In general, most of the results with these units closely parallel those achieved with rotary vacuum filters. They do have advantages in that there is no sludge pickup problem which sometimes occurs with rotary vacuum filters, and they have a lower energy consumption.

These results are satisfactory for many installations and the Continuous Belt Filter Press of this first type or its immediate successor, a two-level unit of the same basic design and concept (see figure 6–9) has in the past 5 years become the most frequently selected dewatering device around the world.

There have been additional developments of the basic principles of the Continuous Belt Filter Press and several third generation units from various companies are now available. In a broad sense these latest improvements may be described as:

1. The addition of some form of continuous mechanical thickening device as the initial stage of a Continuous Belt Filter Press.
2. The addition of additional medium and/or high pressure press sections to the Continuous Belt Filter Press, and variations in the cake shearing mechanisms to obtain additional dewatering.

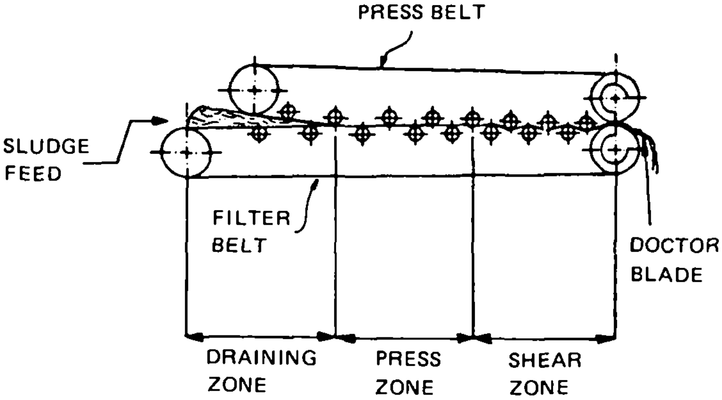


Figure 6–8.—Original concept: continuous belt filter press.

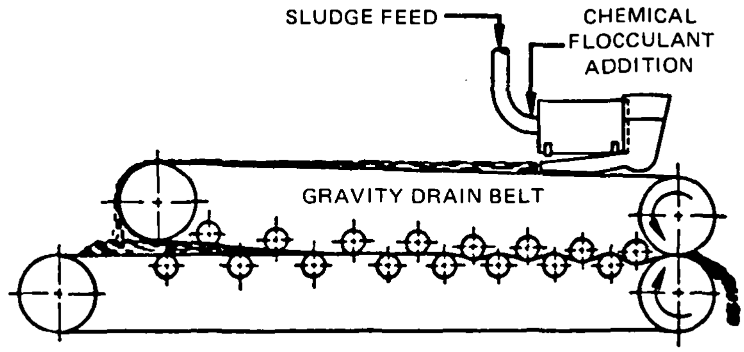


Figure 6–9.—Second generation: continuous belt filter press.

A schematic conceptual drawing of the R. B. Carter Series 31/32 device, the design of same being based on the Klein "S" Press (a unit widely installed around the world) typifies the third generation type unit.

Referring to figure 6–10 and figure 6–11, this device functions as follows:

1. The reactor conditioner (rotating cylindrical screens) removes free draining water, usually increasing sludge solids content from 0.1–0.5 percent to 3–5 percent.
2. The sludge then passes into the first or low pressure zone of the belt press proper with the top belt being solid and the lower one being a sieve belt. Herein further water removal occurs and a sludge mat with significant dimensional stability is forming.

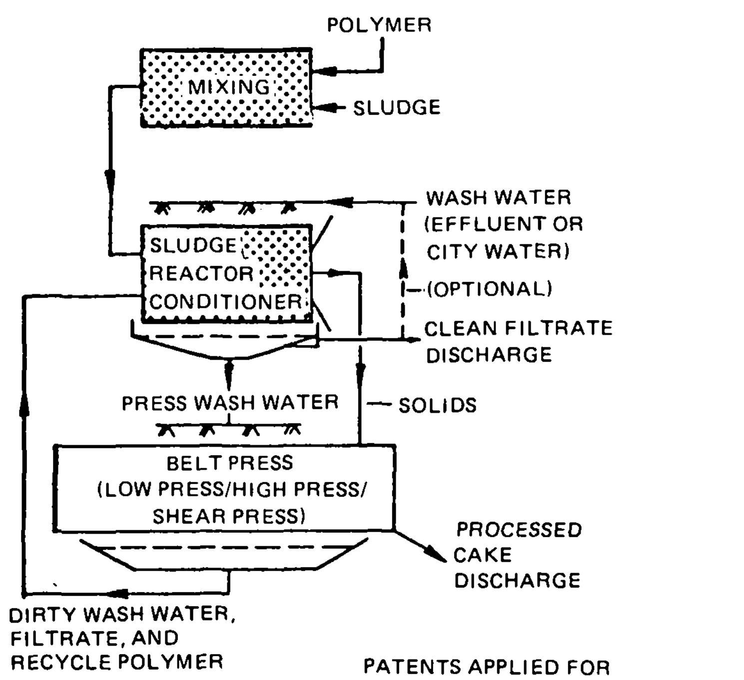


Figure 6–10.—Conceptual schematic: R.B. Carter series 31/32—CBFP.

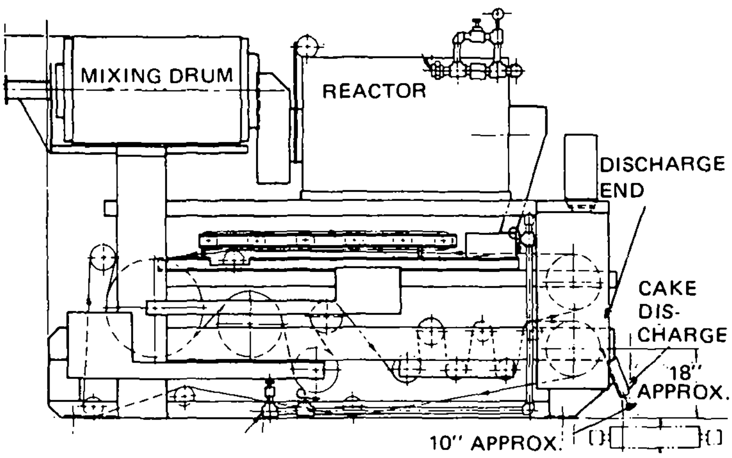


Figure 6–11.—R.B. Carter series 31/32—CBFP.

3. In the second or high pressure zone (4 atmospheres) the sludge is sandwiched between two sieve belts. Large mesh openings are possible because the sludge has developed structural integrity at this point.
4. A serpentine configuration makes up the Shear Zone at the end of the second pressure zone wherein by stretching the belts and sludge cake over smaller rollers, a squeezing action expels more water from the cake.

As will be noted subsequently in more detailed descriptions of each unit, the advanced third generation CBFP's give cake dry solids contents equivalent to those achieved with pressure filters.

In addition to the Carter Series 31/32 device, other suppliers of similar third generation type devices are:

| *Company* | *Unit* |
|---|---|
| Komline Sanderson | Unimat |
| Parkson Co. | Magnum Press |
| Ashbrook Simon Hartley | Winklepress |
| Carborundum | Sludge Belt Filter Press |
| Tait Andritz | SDM |

There are also other Continuous Belt Filter Presses which are more advanced than the original first generation type units. These are also described later.

## Categorization of Continuous Belt Filter Presses

Only units which have at least two phases built into their operation, and which yield cakes which are truly dewatered and dimensionally stable (nonflowable) can logically be classified as Continuous Belt Filter Presses. The Dual Cell Gravity (DCG) Concentrator as supplied by Permutit when used in series with the Permutit multiple roll press (MRP) is a system which performs as a continuous dewatering device in a fashion analogous to the first generation CBFP.

All of the variations start with a gravity drainage zone followed by various combinations of shear and different levels of pressure (or vacuum) applied to the gravity drained cake. Rather than attempting to lump presses of different configuration into rigid categories, each will be described and results listed.

## SMITH AND LOVELESS (S & L) SLUDGE CONCENTRATOR

This device, as described in reference 10, was developed and is marketed by the Smith and Loveless Division of Edodyne. It is essentially a "Gravity-Pressure" filtration unit which uses an endless, variable speed, relatively open mesh filter screen to retain flocculated solids while the bulk of liquid passes through the screen. Solids from the gravity drainage stage pass into the second or pressure stage where three sets of compression rollers further dewater the cake. The pressure increases with each set of rollers. The dewatered sludge falls off the belt into a discharge chute for removal.

The S & L Concentrator is offered in two models of varying size. Typical dewatering capacities claimed are described in table 6–11.

As will be noted this device does not give as dry a cake as some of the other more complicated machines. It has found usage at certain plants which can utilize cake solids levels as shown. The unit uses only 5 horsepower versus a normal 40 horsepower for a rotary vacuum filter.

## PERMUTIT DCG—MRP

This system consists of a dual cell gravity unit followed in series by a multiple roll press. In reference to the schematic cross section of the DCG, this first drainage section forms a plug of fluid sludge in the first fine mesh nylon cell and then the plug is further dewatered in cake form in the second cell (see figures 6–12 and 6–13).

The relatively moist cake from the DCG is conveyed to the MRP, an inclined dual continuous spring loaded belt which further dewaters the sludge cake.

Table 6–11.—S & L sludge concentrator performance estimate

| Type of sludge | Estimated dewatering rate (lb/hr) Model 40 | Model 80 | Polymer dosage lb/ton | Cake solids, percent |
|---|---|---|---|---|
| Anaerobic digester primary | 250 | 500 | 15 | 12 |
| Aerobic digester W.A.S. | 250 | 500 | 10 | 10 |
| W.A.S. | 225 | 450 | 10 | 10 |

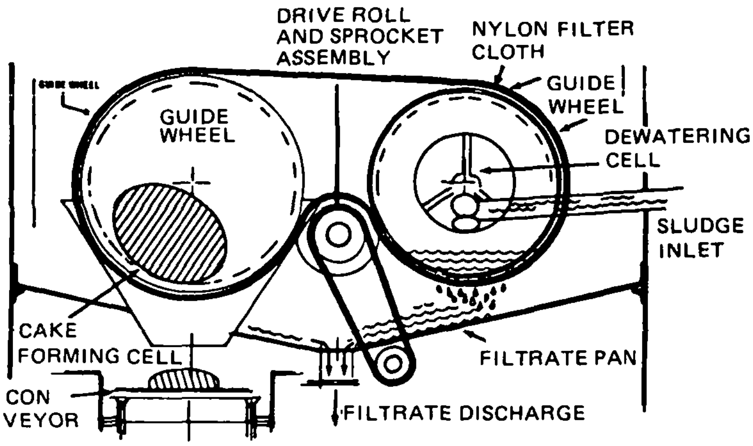


Figure 6–12.—Cross section of a Permutit DCG.

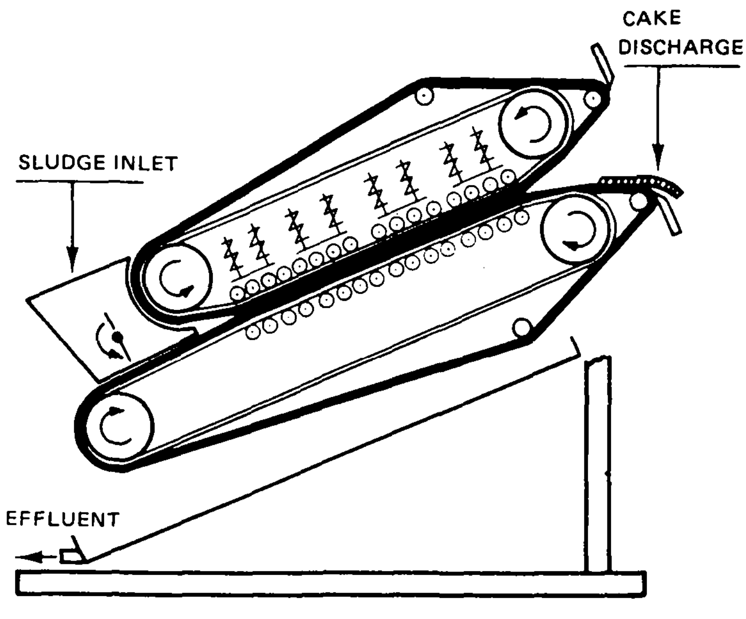


Figure 6–13.—Cross section of a Permutit MRP.

Typical performance on the DCG-MRP (Caldwell, N.J.) indicates dewatering of an anaerobically digested mixture of primary and humus sludge from a feed solids of 4–5 percent yielding a dewatered cake of 15 percent dry solids with polymer costs of $8 to $10 per ton ($9 to $11 per Mg).

The DCG-MRP has worked reasonably well at small plants with noncontinuous dewatering schedules. Some problems have been noted with maintainability of the early units and some modifications are in process.

## INFILCO DEGREMONT FLOC-PRESS

This is a two stage unit of French origin featuring a horizontal belt gravity drainage area on a woven synthetic fiber belt followed by a press section. The partially dewatered cake is sandwiched between the lower belt and a rubber pressure belt (adjustable hydraulic loading) to provide cake solids levels similar to that which is obtained in rotary vacuum filters or centrifuges (see figure 6–14).

There are 46 world-wide Floc-Press installations and there were five in the United States as of January 1976. A notable U.S. installation is at Medford, N.J.[11] At Medford, a 0.9 Mgal/d (0.04 $m^3/s$) contact stabilization plant, a two meter wide Floc-Press replaced an existing rotary vacuum filter which has been shut down. The results are shown in table 6–12.

The horsepower consumption is 6.25 (4.7 kW) for the Floc-Press versus 22 (16.4 kW) for the previously used rotary vacuum filter. The RVF had provided similar cake solids but poorer solids capture. Polyelectrolyte costs are in the $11–15/ton ($12–16/Mg) range. The filter belt is still in excellent condition after almost a year of operation. The wash water rate is 22 gpm (1.4 l/s) at 50 lb/in.$^2$ (3.5 kg/cm$^2$) and plant effluent water is used.

The Floc-Press system includes a mounted sludge conditioning chamber and other auxiliaries such as chemical conditioner and sludge feed systems, conveyors for sludge removal and automated control panels.

Output in pounds per foot of belt width per hour is quoted at 134–268 (200–400 kg/m) for an anaerobically digester mixture of primary and W.A.S. at a feed solids of 3.5 to 9 percent. the Medford, N.J., Floc-Press is 16

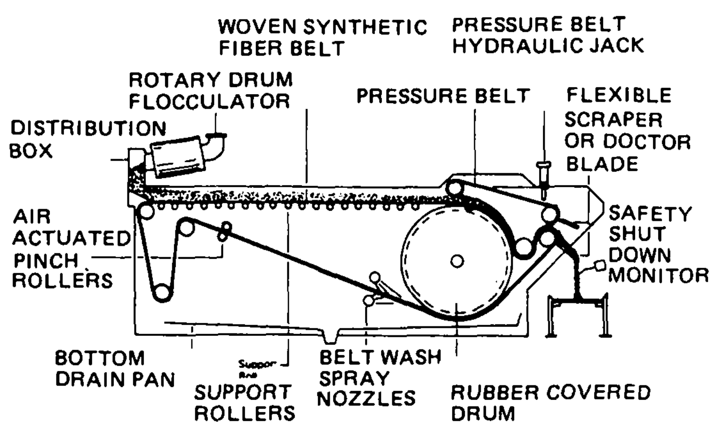


Figure 6–14.—Infilco Degremont Floc-Press.

Table 6–12.—Floc press results—Medford, N.J.

| | Averages |
|---|---|
| Feed solids, percent | 3–4 |
| Cake | 17–19 |
| Filtrate suspended solids (PPM) | 100 |
| Percent solids capture | 98 |

feet 1-1/4 inches long, 10 feet 4-3/8″ wide, and 10 feet 6 inches (4.9 m × 3.1 m × 3.2 m) tall.

The Floc-Press is available in belt widths varying from a nominal 3 feet (0.9 m) to a nominal 10 feet (3 m) with effective belt areas of 32.28 square feet to 96.84 square feet (3–9 $m^2$). For the larger units, only additional width must be provided for.

## PASSAVANT VAC-U-PRESS

This is a German development which features the following:

1. A continuous press utilizing gravity and vacuum drainage followed by a pressure zone.
2. Conditioned sludge is evenly distributed on a moving belt which initially drains by gravity and then by virtue of vacuum boxes beneath the belt.
3. The compression belt is applied on top of sludge on the lower belt to form a sandwich.
4. The two belts are subjected to pressure by going under tension around large dewatering cylinders. Pressure is then applied to alternating sides of the belt by smaller pressure rolls.
5. Dewatered sludge is discharged and belts are continuously back-washed.
6. The Vac-U-Press is enclosed in a fiberglass reinforced polyester housing to control noise and odor.

Typical sizing data are as shown in table 6–13.

There are five U.S. installations of the Vac-U-Press, all of the BFP-200 model. Indications are that it gives a dewatered cake slightly drier than a rotary vacuum filter. A mobile test unit is available for rental.

## TAIT ANDRITZ SDM and SDM-SM

Andritz, an Austrian equipment firm, first developed a continuous double belt filter dewatering device for use on various industrial sludges. In the past two years Tait Andritz of Lubbock, Tex., has sold and installed 43 of these devices at 28 total U.S. locations for dewatering of various industrial and municipal sludges. The 1977 world-wide installation list shows 68 locations where these devices are in use. Twenty of these locations are on municipal sludges. The industrial installations are in some cases on straight 100 percent biomass sludges.

The dewatering in the Tait Andritz unit(s) is achieved

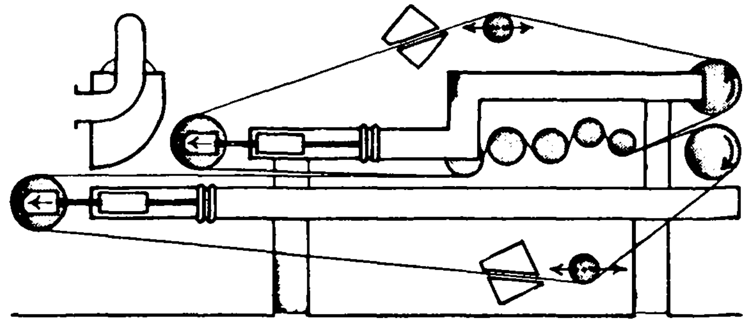

Figure 6–15.—Tait Andritz—SDM-SM model.

by passage of the sludge through a gravity dewatering zone, into a wedge zone for pressure dewatering, followed by higher pressure dewatering in a module zone. The module zone can be either an S configuration (offset rolls), or a press configuration (pressure loaded rollers).

Main design features are: variable speed drive for belts and conditioning drum; pneumatic belt tensioning and pressure adjustment during operation; pneumatic belt tracking; and in the industrial SDM model, use the endless belts.

Figure 6–15 shows the SDM-SM model (seamed belts) designed for municipal operation where unattended round-the-clock operation is not necessary.

Table 6–14 summarizes reported operating results.

The results shown in table 6–14 tend to indicate that the Tait Andritz CBFP's will normally produce a cake solids content somewhat higher than that obtainable in a rotary vacuum filter. Further, more definitive results on the two versions (either the "Press Module" or the "S" Module equipped) of the basic device will be forthcoming during 1978. In this vein, it is understood that Burlington, Wis. (an installation discussed later) has recently ordered several units.

The Tait Andritz SMD device (Industrial) has an excellent performance record (ease of maintenance, etc.) in dewatering biological and mixed sludges in the paper industry.[12,13]

Data on the size of the three SDM-SM models available are shown in table 6–15.

Table 6–13.—Passavant Vac-U-Press—sizing data

| Model Number | Belt width | Length (ft/in.) | Width (ft/in.) | Height (ft/in.) | Drive motor (hp) | Active belt area ($lb/ft^2$) | Nominal capacity (gal/hr) |
|---|---|---|---|---|---|---|---|
| BFP075 | 26-1/2 | 14–9 | 4–1 | 5–3 | 1.5 | 90 | 1,500 |
| BFP125 | 43-1/2 | 14–9 | 5–8 | 5–3 | 3 | 150 | 2,500 |
| BFP200 | 72-1/2 | 14–9 | 8–2 | 5–3 | 3 | 250 | 4,200 |

Table 6–14.—Tait Andritz—SDM-SM results

| Type sludge | Percent dry solids | | Throughput[a] | | Polymer cost ($/ton dry solids) |
|---|---|---|---|---|---|
| | Feed | Cake | gal/min | Dry solids (lb/hr) | |
| Raw primary | 5–7 | 22–26 | 10–14 | 300–500 | 4–7 |
| Primary and W.A.S. | 3–5 | 20–25 | 15–20 | 200–350 | 4–8 |
| Unox ext. aer. | 1–2 | 18–23 | 20–25 | 200–250 | 8–10 |

[a]Per 20 inches of working belt width.

Table 6–15.—Tait Andritz SDM-SM—machine sizing data

| Size and type | Working belt width (in.) | Overall dimensions | | | Weight (lb) | Conn H.P. load | Belt spray consumption (gal/min) |
|---|---|---|---|---|---|---|---|
| | | Length (in.) | Width (in.) | Height[a] (in.) | | | |
| SDM 40 | 40 | 152-1/2 | 75 | 75 | 5,513 | 3-1/2 | 18–24 |
| SDM 60 | 60 | 186 | 114 | 83 | 14,333 | 5-3/4 | 30–37 |
| SDM 80 | 80 | 186 | 134 | 83 | 17,640 | 5-3/4 | 35–45 |

[a]Height will vary according to drive system used.

## ASHBROOK SIMON-HARTLEY WINKLEPRESS

The Winklepress was developed by Gebr. Bellmer KG. of Germany. Simon-Hartley of the United Kingdom markets U.S. units through a subsidiary, Ashbrook Simon-Hartley of Houston, Tex.

Figure 6–16 is a schematic conceptual drawing which shows that the device employs two endless synthetic fiber mesh sieve belts to convey and dewater conditioned sludge. After an initial gravity sandwich drainage stage, the primary belt meets the second belt and forms a vertical sandwich drainage section. The two belts, which are under tension, then carry the sludge along an arrangement of staggered rollers where multiple shear force action areas squeeze out remaining free water. The sieve belts are continuously washed.

While there are a number of operational installations in Europe, as of November 1, 1977 none of the U.S. installations under construction had started operation. (See tables 6–16 and 6–17.)

## KOMLINE SANDERSON UNIMAT $GM_2H$-7 CONTINUOUS BFP

Komline Sanderson manufactures its version of the German Unimat under license from Mull-Abwasser-Transportanlagen-GMBH, Elversberg, West Germany.

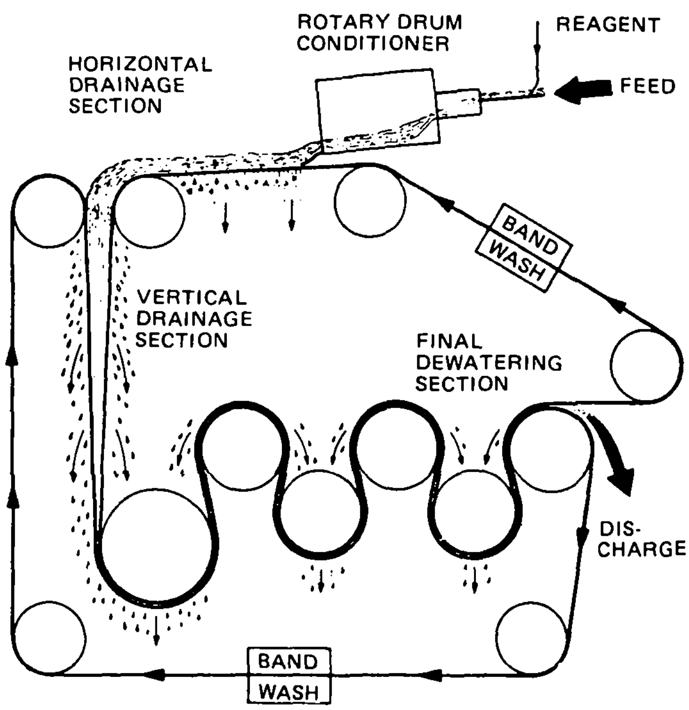


Figure 6–16.—Schematic of an Ashbrook Simon-Hartley Winklepress.

Table 6–16.—Winklepress test results (from supplier)

| | Dry solids | | Filtrate (mg/l) | Polymers (kg/m³) | Capacity feed | |
|---|---|---|---|---|---|---|
| | Feed | Cake | | | m³/h meter | gal/min |
| Digested primary and humus | 3.8 | 36.2 | 85 | 0.182 | 7.5 | 33.0 |
| | 5.7 | 36.3 | 95 | 0.165 | 6.5 | 28.6 |
| Digested primary and W.A.S. | 3.5 | 36.3 | 90 | 0.165 | 7.5 | 33.0 |
| | 4.8 | 38.5 | 75 | 0.182 | 7.5 | 33.0 |

Table 6–17.—Winklepress size and capacity data

| Winklepress size | Input width | | Nominal capacity of digested sludge | |
|---|---|---|---|---|
| | mm | inches | m³/h | gal/min |
| 0 | 200–300 | 8–12 | 2–3 | 8.8–13 |
| 1 | 500–800 | 20–32 | 5–8 | ,22–35 |
| 2 | 1,000–1,300 | 39–51 | 10–13 | ,44–57 |
| 3 | 1,500–1,800 | 59–71 | 15–18 | ,61–79 |
| 4 | 2,000–2,300 | 79–91 | 20–23 | ,88–101 |

The most advanced model of the modularized Unimat (figure 6–17) which is designed for maximum cake dryness and throughput is the $GM_2H$-7. This press consists of four stages:

1. Gravity drainage (actually a thickening stage)
2. A mild pressure stage
3. A medium pressure stage
4. A high pressure stage

The initial gravity drainage stage is a continuous belt of pockets which are formed by folding a rectangular piece of cloth. This is a separate belt. After thickening in this first stage the sludge dumps into a different belt which moves over a gravity drainage tray prior to dumping onto another belt on a succeeding tray (and a dif-

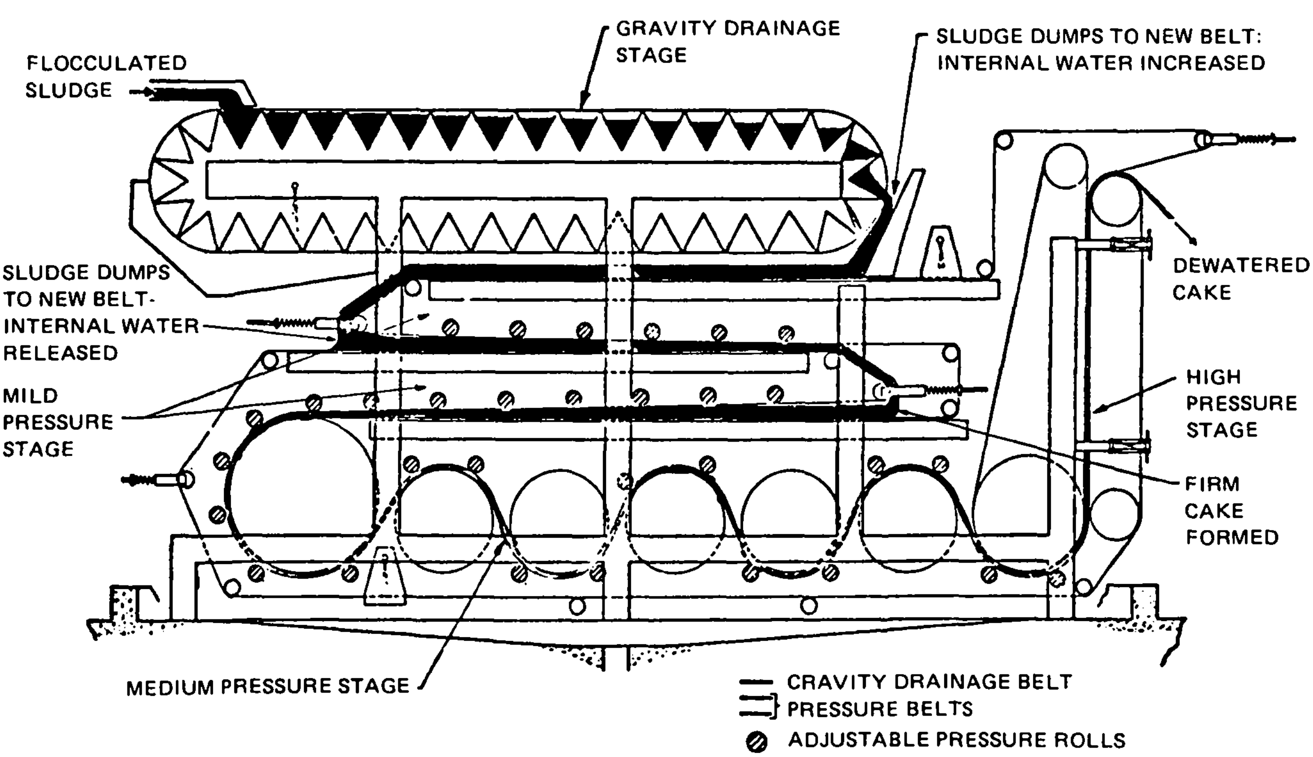


Figure 6–17.—Komline Sanderson Unimat $GM_2H$-7.

ferent belt) where a small amount of pressure is added by small diameter rollers. Then it is moved to the third tray of the mild pressure section (and back on the original pressure belt) and subjected to slightly more pressure before going into the medium pressure stage. All the rollers in the medium pressure stage are adjustable for pressure optimization. While passing over the medium pressure rolls, the cake sandwich between the belts is flexed from one side to the other. Each of the large diameter drums has smaller diameter rolls which apply pressure as the sandwich passes over the drums. Every other roll is perforated for water removal. Pressure is applied to the cake by tension on the belts as the belts go around the drums and by the small diameter rollers. The belt tension is, however, relatively low and all synthetic media is used instead of stainless steel in the long axis.

The cake now goes to the high pressure stage which can be thought of as two caterpillar tractors standing upright with the tracks butting together. As in the medium pressure section the pressure is adjustable through springs.

In applications where a very high dry solids in the cake is not imperative, the unit is available without the high pressure section.

In addition to the previously mentioned nomenclature and model system the Unimat series is available in three models:

Model S Gravity stage
Model SM Gravity and medium pressure stages
Model SMH Gravity, medium and high pressure stages

Table 6–18.—Active filtration surface areas and retention times

| Machine model | Machine width (meter) | Active filtration surface area ($ft^3$) | | Retention time (min) | |
|---|---|---|---|---|---|
| | | *S* | *L* | *S* | *L* |
| S | 1 | 68 | 104 | 1.2 to 6 | 2 to 9 |
| | 2 | 136 | 208 | | |
| | 3 | 204 | 312 | | |
| | | *5 roll* | *7 roll* | *5 roll* | *7 roll* |
| M | 1 | 101 | 190 | 5 to 19 | 10 to 36 |
| | 2 | 203 | 380 | | |
| | 3 | 305 | 570 | | |
| | | *ALL* | | *ALL* | |
| H | 1 | 32.9 | | 2 to 6 | |
| | 2 | 65.6 | | | |
| | 3 | 98.4 | | | |

Note: When using 2 or more sections, the retention time and active surface areas are cumulative.



Table 6–18 lists the design features of this series.

There were 69 European locations employing the Unimat as of November 1976, with practically all of them processing municipal sludges of some type, including straight 100 percent biomass.

Table 6–19 lists reported results.

While at the time of writing this, no Unimat systems are yet operating in the United States, 16 units have been sold and some will be operative by early 1978.

A mobile test unit is available and considerable U.S. test work was carried out on site during 1977.

### Performance of Unimat on Washington, D.C., Mixed Sludge

At Blue Plains the Unimat $GM_2H$-7 dewatered a sludge mixture of 1 part primary plus 2 parts W.A.S. to a dry solids content of 27 to 33 percent at rates of 196 to 206 lb/hr/ft (292 to 307 kg/hr/m) width. Polymer costs were mostly between $8.76 to $9.20 per dry ton ($9.66–$10.14/Mg) with a solids capture of 95–98 percent. On the existing rotary vacuum filters a total dry cake solids of 22–24 percent (including solids resulting from use of 5–7 percent ferric chloride and 15–20 percent lime) is normally obtained. Because of the large variation of the sludge quality, the lime dosage for the rotary vacuum filters reaches 30–40 percent on occasion.

At Blue Plains, the dewatered vacuum filter cake was fed to the $M_2H$ sections of the Unimat and the cake solids were increased to 37–40 percent at a feed rate of 366 lb/hr/ft (545 kg/hr/m) width with no auxiliary conditioner dosage.

### Performance of Unimat on Columbus, Ohio, Southerly Plant Sludge

At Columbus Southerly plant, the anaerobically digested mixture of primary and W.A.S. was dewatered to a cake solids content of 36–39 percent at a rate of 228–305 lb/hr/ft (341–455 kg/hr/m) width. Solids capture was 90–95 percent and polymer costs $8–$14/ton ($9–$15/Mg). Feed solids were 3–4 percent dry solids. Thus an autogenous cake is feasible with this difficult sludge.

It is quite apparent that the K.S. Unimat press is one of the CBFP's newly introduced into the U.S. from Germany which has the capability to effectively dewater mixtures of primary and W.A.S. sludges to a dry solids content high enough to be in the autogenous incineration range.

## PARKSON MAGNUM PRESS

This device, of Swedish origin, is manufactured and sold in the United States by the Parkson Corporation of Ft. Lauderdale, Fla.

The Magnum Press is an advanced or third generation type CBFP designed to maximize dry solids content of dewatered cake. The Magnum Press has three stages and can best be described by reference to the cross sectional side-view of figure 6–18.

Table 6–19.—Dry solids of cake and polymer dosage

| Unimat | Model S | Model SM | Model SMH | |
|---|---|---|---|---|
| Type of sludge feed conc. (percent D.S.) | After gravity stage (percent D.S.) | After gravity and medium pressure (percent D.S.) | After gravity and medium and high pressure (percent D.S.) | Typical polymer dosage (lbs/ton D.S.) |
| Fresh-primary—raw (4–6 percent) | 12–18 | 25–35 | 30–45 | 6.0–8.5 |
| Fresh primary and trickling filter (3–5 percent) | 10–15 | 22–32 | 28–40 | 6.0–10.0 |
| Fresh primary and activated (3–5 percent) | 10–15 | 17–27 | 25–35 | 6.0–10.0 |
| Anaerobically digested primary and activated (4–9 percent) | 14–24 | 25–35 | 30–45 | 5.0–8.5 |
| Activated—100 percent W.A.S. (0.5–1.0 percent) | 8–12 | 17–20 | 17–23 | 7.0–10.0 |

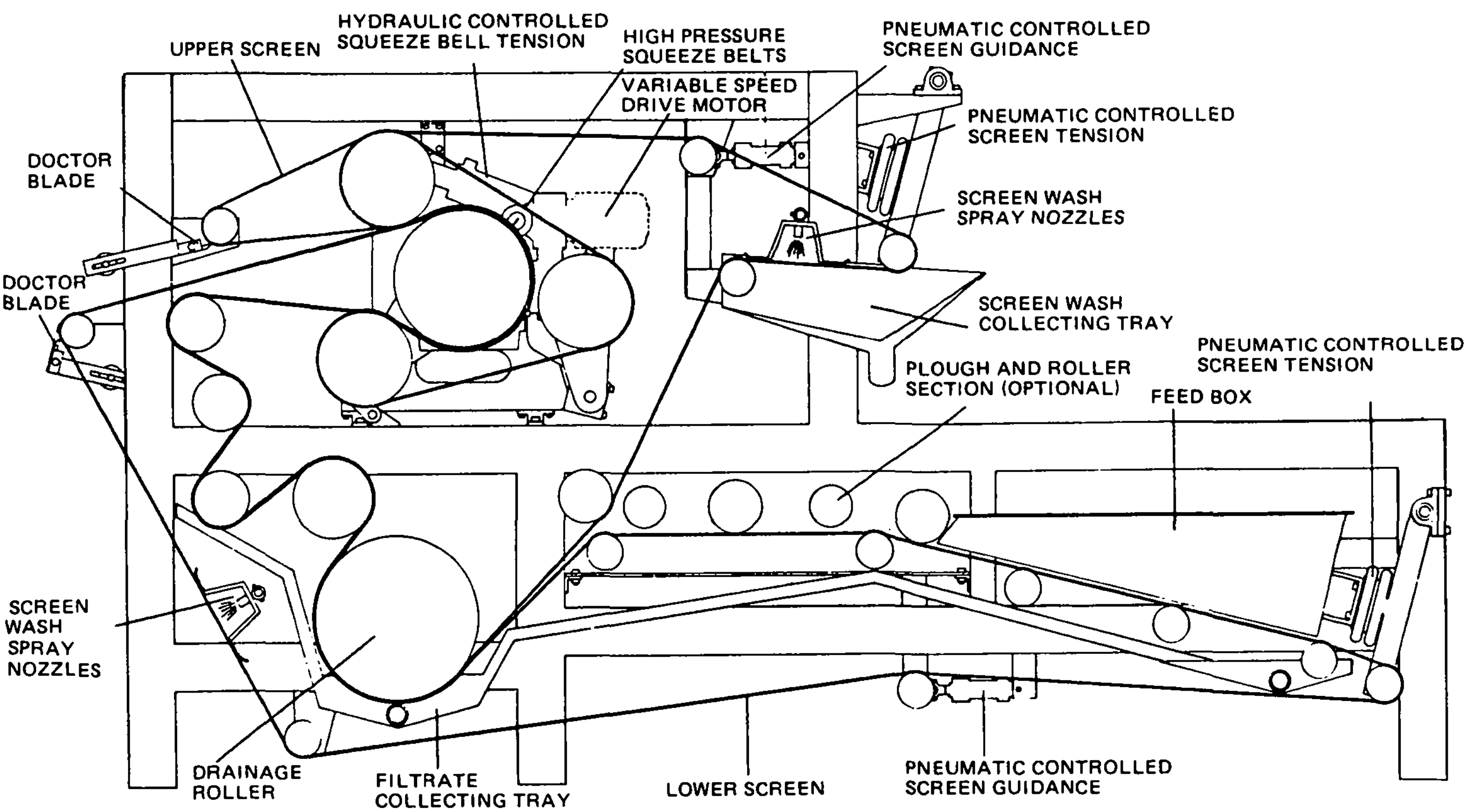


Figure 6–18.—Cross section of a Parkson Magnum Press.

The initial stage is a unique gravity drainage section. In addition to normal dewatering occurring by gravity from a properly conditioned sludge, the sludge can be subjected to a light pressure involved by rollers and be turned by plows (both optional). The partially formed cake then proceeds to the low pressure stage where the second polyester screen belt comes into play on the top forming a sandwich that is fed into the second or low pressure stage. In the low pressure stage perforated press rolls of decreasing diameter subject the cake to continuously increasing pressures. In the last or high pressure stage the cake is subjected to very high pressure that is adjustable, depending on the application. The high pressure is generated by a series of 1 inch wide flat belts that press the screens against a perforated roll uniformly from side to side. This feature allows

Table 6–20.—Magnum press size data

| Model | Screen width (nominal) | Weight (tons) | Overall dimensions | | | Screen wash water flow rate @ 100 gal |
|---|---|---|---|---|---|---|
| | | | A-width | B-height | C-length | |
| MP-20 | 20″ | 3.8 | 4′ | 7′–9″ | 14′–10″ | 12 gal/min |
| MP-40 | 40″ | 4.4 | 5′–8″ | 7′–9″ | 14′–10″ | 24 gal/min |
| MP-60 | 60″ | 4.8 | 7′–4″ | 7′–9″ | 14′–10″ | 36 gal/min |
| MP-80 | 80″ | 6.0 | 9′ | 7′–9″ | 14′–10″ | 48 gal/min |

the sludge to be subjected to high pressure for a long period of time without producing an excessive load on the screens. The pressure is adjustable through the use of two hydraulic cylinders.

This final high pressure stage of the Magnum Press can also be employed in a modular fashion to further dewater filter cake from existing Rotary Vacuum Filter installation.

The Parkson Magnum Press is available in four sizes as shown in table 6–20.

As of December 1977, nineteen Magnum Presses had been sold worldwide. There are seven Japanese installations, nine in Europe, and three in the United States. The first U.S. unit (at Mobil Oil Co.) processing straight excess biological sludge is just now commencing operation.

Parkson has a mobile Magnum Press and a smaller pilot unit, both of which have been used to carry out on-site tests at various U.S. locations.

## Performance of Magnum Press at Washington, D.C.

A 0.25 meter pilot unit was evaluated on the various sludges at Blue Plains plant. The following two figures show the results obtained with various mixtures of primary and excess activated sludges (including phosphorus removal sludges resulting from iron salt use).

In assessing results of dewatering work at Blue Plains it is important to note the following:

1. The normal mix is 32 percent raw primary/68 percent raw secondary sludges (on a weight percent dry solids basis). The primary is gravity thickened to 9.5 percent and the secondary is DAF thickened to 5.5 percent. The resulting 6.8 percent solids mix is filtered on RVF's to about 18 percent (without lime).
2. The Blue Plains plant has an abnormally large amount of a difficult to process excess activated sludge due primarily to the use of a high rate activated sludge biological treatment system. This system was apparently chosen because of certain site and capacity constraints.

As can be seen in figure 6–19, the Magnum Press produced a dewatered cake of 30 percent dry solids

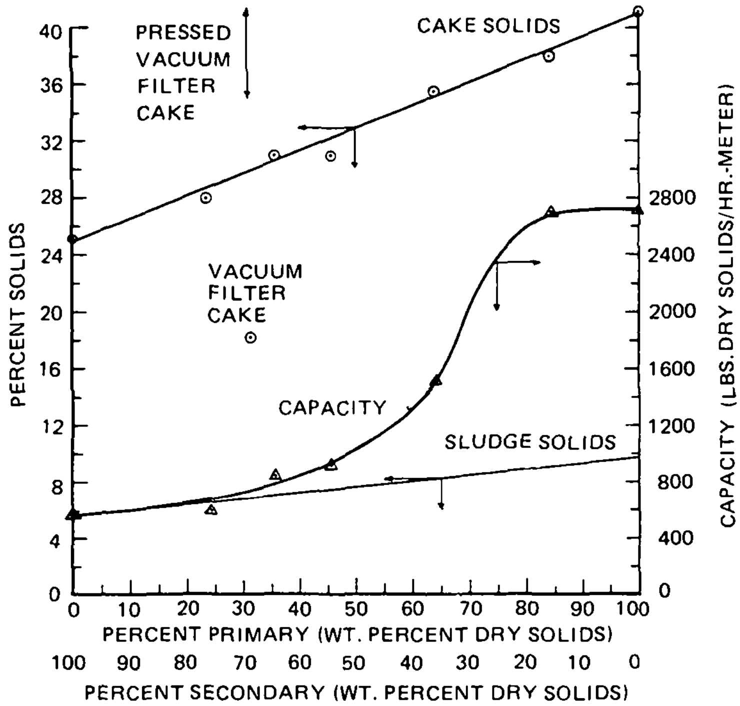


Figure 6–19.—Magnum Press results, Blue Plains.

content at a rate of 244 lb/hr/ft (364 kg/hr/m) belt width.

It should also be noted that a straight interpolation of the data in figure 6–20 indicates that at a more normal sludge ratio of 60 percent primary and 40 percent secondary, even with the high rate W.A.S., the production rate would be 17 percent greater and the cake solids would be 34 percent. As shown in figure 6–20, polymer dosages varied from 5.5 to 1.6 pounds per ton (2.8 to 0.8 kg/Mg) of dry solids and solids recoveries varied from 95 to 98 percent.

The Magnum Press was also tested for dewatering the filter cake from the existing RVF's. Cake solids of 35–42 percent were obtained at rates of 244 to 853 lb/hr/ft (364 to 1273 kg/hr/m) belt width. There is mechanical development work required to design equipment to transfer the filter cake to such a press.

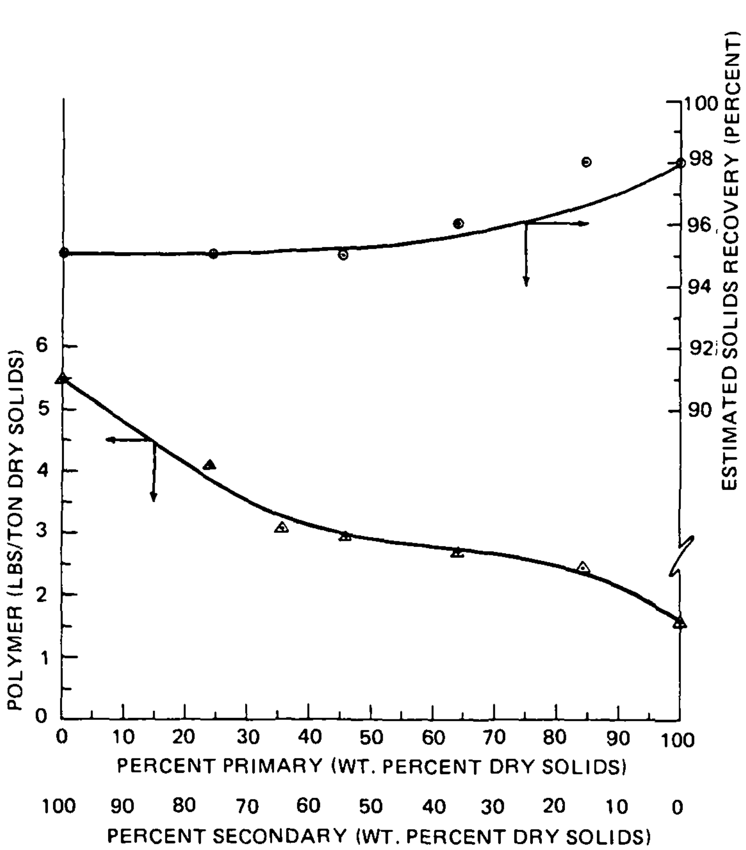


Figure 6–20.—Magnum Press results II, Blue Plains.

## Magnum Press Performance at Los Angeles/Orange County Metropolitan Area (LAOMA)

The Magnum Press mobile unit was evaluated on several mixtures of the sludges being studied in this major research and development project.

While the results in table 6–21 are impressive and may well be acceptable for the system, it is also apparent that the dewatering devices' performance is penalized by attempting to dewater an unthickened sludge. It is strongly suspected that if the LAOMA sludges were thickened a much higher capacity and cake solids would be realized, in addition to being operable at a much lower polymer dosage.

## Magnum Press Performance—Other Locations

A bench scale Magnum Press has been evaluated at various other locations in table 6–22.

It is significant to note that the Magnum Press will function with inorganic conditioning agents to extend the flexibility of the unit and to reduce polymer costs.

## CARBORUNDUM SLUDGE BELT FILTER PRESS

Carborundums' Pollution Control Division at Knoxville, Tenn., manufactures and sells a unit called the Sludge Belt Filter Press (SBFP). This unit is based on the design of Rittershaus and Blecher of Germany who developed the "Dreibandpresse."

The Carborundum unit incorporates two unique features: stainless steel wire supported belts and oscillating pressure rollers.

As can be seen in figure 6–21, the gravity drainage section of the SBFP includes two phases involving a dumping of the partially drained sludge from the initial belt onto a second drainage belt prior to the incidence of the upper sandwiching belt. The two belt cake sandwich then proceeds around a large diameter roll into a further pressurizing section involving smaller diameter offset pressure rollers in a two level configuration. Thus, in effect, the Carborundum SBFP has a two stage gravity drainage section plus two additional pressureshear stages to successively expose the cake to increasing degrees of shear and pressure.

Carborundum is also bringing out a newer model with a "Pre-Concentrator" stage in the same vein as the Unimat and R. B. Carter Series 31/32 devices.

The current Carborundum SBFP is available in 2 models. Table 6–23 shows the dimensions.

This unit was introduced into the United States in 1977 so no U.S. commercial scale operating data are yet available. A pilot unit is available for testing and the supplier quotes the results as shown in table 6–24.

Additional field U.S. results are now available from

Table 6–21.—Performance of magnum press—Los Angeles/Orange County metropolitan area

| Sludge mixture (digested mix) | Dry solids, percent | | Capacity-dry solids (lb/hr/m) | Polymer ($/ton dry solids) | Percent solids recovery |
|---|---|---|---|---|---|
| | Feed | Cake | | | |
| 70 Prim–30 W.A.S. | 1.8 | 29 | 360 | 12.60 | 96 |
| 30 Prim–70 W.A.S. | 2.1 | 21 | 320 | 21.40 | 88 |

Table 6–22.—Performance of Magnum Press—various sludges[a]

| Location | Sludge mixture | Dry solids, percent Feed | Cake | Capacity-dry solids (lb/hr/m) | Flocculant ($/ton dry solids) | Percent solids recovery |
|---|---|---|---|---|---|---|
| Blue Lake, St. Paul, Minn. .. | 45–Prim.[b] 55–W.A.S. | 5.3 | 35 | 1,260 | 14 | 98 |
| Lake Charles, La. ............ | Prim + W.A.S. | 2.9 | 29–34 | 580 | 12 | 95 |
| Richardson, Tex. .............. | Digested prim. + W.A.S. + alum | 4.1 | 26–27 | 615 | [c]11 | 95 |
| Industry ........................ | W.A.S. | 3.5 | [d]22–23 | 500 | [d]17 | 95 |

[a]All results from 0.25 meter bench scale press.
[b]Concentrations by volume.
[c]Costs using 75 lb/ton $FeCl_3$ plus 5 lb/ton polymer: straight polymer = $16/ton.
[d]Values shown are for 100% polymer usage: use of 30–55 lbs/ton $FeCl_3$ will increase cake solids to *net* of 27% at slightly lower capacity.

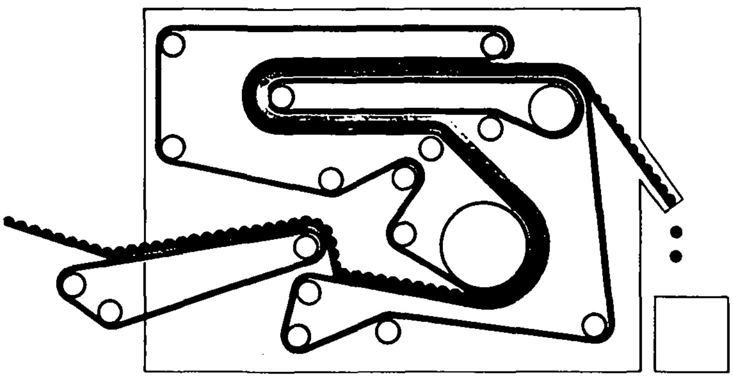

Figure 6–21.—Carborundum sludge belt filter press.

Table 6–23.—Carborundum SBFP

| Model | Belt width (in.) | Approximate overall dimensions (inches) Length | Height | Width |
|---|---|---|---|---|
| 135 .................................... | 39 | 160 | 96 | 69 |
| 215 .................................... | 70 | 160 | 96 | 100 |

Carborundum and German full-scale installations have been in operation for several years.

## R. B. CARTER SERIES 30 PRESSES

R. B. Carter of Hackensack, N.J., is the U.S. licensee of Klein of Germany, the developers of three successive generations of continuous belt filter presses, each of increasing capability in either capacity or cake solids content realized.

Table 6–24.—Carborundum SBFP results

| Type sludge | Capacity (gal/hr) | Feed solids (percent) | Cake solids (percent) | Polymer cost ($/ton dry solids) |
|---|---|---|---|---|
| Primary + W.A.S. ....... | 900 | 4–6 | 34–37 | 9 |
| Anaerobically digested primary + W.A.S. ..... | 1,300 | 4–9 | 26–40 | 10 |
| W.A.S. .................. | 1,100 | 4 | 16–20 | 11 |

The original single level Klein device which was introduced in Germany in about 1969, the Carter Series 30 (a two level unit), and the lastest multistage unit, the Carter Series 31/32 CBFP (based on the Klein "S" press) were described in a preceding section dealing with the evolution of the CBFP. The early single level device has been superseded by the two level Series 30 and the multistaged Series 31/32.

## R. B. Carter Series 30 Installations, Dimensions and Results

As of July 1976 there were 21 U.S. installations of the Carter Series 30 CBFP that were either operating or were on order. The 21 installations involved 36 units. Of these installations, 8 were for industrial sludges and 13 municipal.

The series 30 units are available in 3 sizes as shown in table 6–25.

Table 6–25.—Carter series 30—overall dimensions

| Model | Width (inches) | Weight (lbs) |
|---|---|---|
| 5/30 | 53 | 2,500 |
| 10/30 | 73 | 3,500 |
| 15/30 | 93 | 4,500 |

Table 6–26.—Performance data—Carter series 30 CBFP

| Type sludge | Solids content (percent) | | Capacity (lbs/hr/sq.ft) | Polymer ($/ton dry solids) |
|---|---|---|---|---|
| | Feed | Cake | | |
| Primary + W.A.S. | 4–5 | 20–30 | 6.5–12 | 4–8 |
| Anaerobically digested primary + W.A.S. | 6–8 | 20–30 | 10–20 | 4–8 |
| Extended aeration (no primary treat) | 2–4 | 16–24 | 6–10 | 2–6 |

The Series 30 is typically about 12 feet long and five feet tall. Quoted typical results for the Carter Series 30 model are shown in table 6–26.

A mobile pilot unit of the Series 30 has been used in onsite test work.

## Performance of a CBFP of the Carter Series 30 Type in the U.K.

In addition to the quoted typical results above additional insights into the capabilities of the Carter Series 30 units can be gained by study of references 14 and 15. The latter reference is an exhaustive study by the U.K. Department of the Environment (D.O.E.) on an installation of the British version of the first generation Carter type press. This study was carried out over many months by the D.O.E., an agency of the government, at Lenham Works in East Kent.

Different mixtures of sludges were processed to determine applicability of the single level first generation CBFP, including operability, maintainability, and all cost factors as well as dewatering capacity.

Typical results are shown in table 6–27.

As will be noted the normal mixed sludge is not a difficult one and results were essentially equivalent to dewatering with an RVF. However, it is doubtful that an RVF would have achieved results on straight secondary sludge similar to those shown.

The Lenham plant is a small plant designed to treat a dry weather flow of 0.11 Mgal/d (0.005 $m^3/s$) and actually processing about one half of design flow. The plant includes primary, trickling filter and activated sludge operation. Though the normal sludge mixture is a relatively easy to process material, the performance of the first generation CBFP was viewed as highly successful.

The cost analysis (table 6–28) showed a total operating and capital cost of $65.50 per ton ($72.20/Mg) of dry solids dewatered. Maintenance costs were low.

Table 6–27.—Single level press—R. B. Carter type, Lenham Works, East Kent, U.K.

| Type sludge | Dry solids, percent | | Capacity[a] (lbs dry solids/hr) | Polymer ($/ton) | Percent solids capture |
|---|---|---|---|---|---|
| | Feed | Cake | | | |
| Primary + humus + W.A.S. | 4.5 | 22 | 72 | 5.64 | 96–99 |
| Straight humus | 4.5 | 18 | 49 | 8.00 | 96 |

[a]0.5 meter belt width × 3.0 meter length—Wm. Jones, Chem. Eng. Ltd.

Table 6–28.—Lenham Works—cost analysis, first generation CBFP

| Item | $/ton dry solids |
|---|---|
| Polymer | 4.90 |
| Wash water | 1.94 |
| Power | 0.66 |
| Operating labor (inc. super.) | 12.00 |
| Total operating | 19.50 |
| Capital costs | 46.00 |
| Total (ex. maint.)[a] | 65.50 |

[a]Maintenance estimate + 3/4 hour/1,000 hours operation.

## Performance of an R. B. Carter Series 30 CBFP—Hutchinson, Minn.

At Hutchinson, Minn., a Series 30 Carter CBFP has been operating for many months on a municipal sludge from an activated sludge plant. Figure 6–22 is a photo of the unit.

At Huntchinson, the waste activated sludge is fed to the CBFP at a solids concentration of 1–1.5 percent resulting in a cake solids content of 13–15 percent and dry solids throughput of 340 pounds per hour (155 kg/hr). While this performance is satisfactory it could be greatly improved by prethickening to a solids content more logical for maximum dewatering capability.

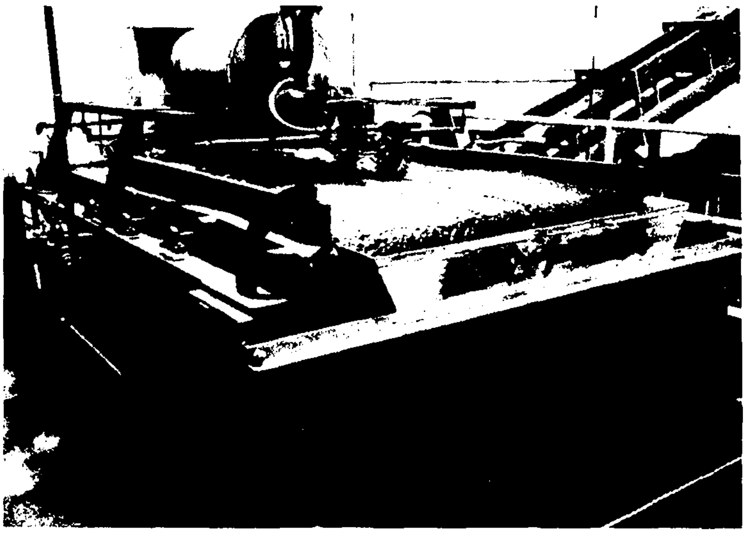

Figure 6–22.—Carter series 30 CBFP.

## R.B. CARTER SERIES 31 CBFP

The basic design characteristics of this unit have been delineated in ealier sections. Essentially it consists of an initial "Reactor Conditioner" system which performs the dual function of conditioning and prethickening followed by two successive pressure zones and a shear zone under pressure.

The Series 31 device also comes in 3 sizes, 5/31, 10/31, and 15/31 which differ in widths. The largest unit, the 15/31, is designed for a nominal feed of 85 gal/min (5.4 l/s) of typically a 5 percent mixed sludge. Complete systems, including the chemical feed system, pumps, controls and erection costs are usually priced at slightly less than $2,000/gal/min ($31,160/l/s) or $170,000 for an 85 gal/min (5.4 l/s) Series 15/31 unit. Solids capture in the Series 31 normally averages 95 percent plus. Connected electrical power, including sludge pumps and conditioner system pumps totals not more than 15 horsepower (11.2 kW).

Sizing of a building or space for a two unit Series 31 system, including polymer preparation system, and conveyor sludge removal system indicates a floor space requirement of about 36 feet by 18 feet (11.0 m×5.5 m). Height requirement is 13 feet 6 inches (4.1 m) minimum.

While there are quite a few operating installations of the Series 31 type unit (Kleins or Wm. Jones "S" Press) around the world, U.S. commercial units were just coming on stream during 1977.

### Performance of R. B. Carter Series 31 CBFP at Hamilton, Ontario

The Carter Series 31 mobile pilot unit has been tested at several North American locations including Hamilton, Ontario, among others.

On a digested mixed primary and W.A.S. sludge at Hamilton, a 27 percent dry solids cake was obtained which compared very favorably with a 16 percent cake being obtained at the same time on the existing Rotary Vacuum Filters. Hamilton was experiencing some problem with fines recirculation and accumulation within the system at the time and no doubt even more favorable results would be realized in a situation with normal sludge conditions.

### Performance of R. B. Carter Series 31 CBFP at Parkersburg, W. VA.

At the Borg Warner Co., two 15/31 Carter units are dewatering a pure excess biological sludge. Feed solids are 0.5 to 2.0 percent with a cake solids content of 25–33 percent. Capacity averages 1500 pounds (682 kg) of dry solids per hour per machine.

### Performance of R. B. Carter Series 31 CBFP at Scituate, Mass.

A Carter Series 31 unit equipped with a Reactor-Thickener was evaluated on the difficult aerobically digested extended aeration sludge at the Scituate, Mass. plant. Results are shown in table 6–29.

In a cost comparison, the engineers involved estimated that a production level of 3 dry tons (2.7 Mg) per day for a 5-day week either 2 Carter Series 31 CBFP's (60 inches wide) with Reactor-Thickener first stages: or two 250 ft$^2$ (23 m$^2$) DAF units plus two 200 ft$^2$ (19 m$^2$) RVF's would be required. Equipment costs for the CBFP option were estimated at $222,000 and for the second option at $425,000. Horsepower requirements were estimated at 26 hp (19 kW) and 200 hp (149 kW) respectively for the two systems.

## DESIGN EXAMPLE—CONTINUOUS BELT FILTER PRESS

### Basic Assumptions

These assumptions are identical to those used in the example for design of a Rotary Vacuum Filter System:

1. Anaerobically digested mixture of primary and W.A.S. at 4 percent solids content, 60 percent primary and 40 percent W.A.S.

Table 6–29.—Carter CBFP—Model 5/31, aerobically digested extended aeration sludge scituate, Mass.

| Test | Percent dry solids | | Sludge feed (lbs/dry solids/hr) | Solids capture, percent | Polymer cost ($/ton dry solids) |
|---|---|---|---|---|---|
| | Feed | Cake | | | |
| 1 | 2 | 18 | 88 | 91 | [a]26 |
| 2 | 3 | 16 | 255 | 98 | [b]11 |

[a]Cationic polymer A used.
[b]Cationic polymer B used.

2. Ultimate disposal by hauling to either a sanitary landfill, or to farmland, composting or other horticultural use.
3. Equilibrium sludge removal rate of 2.5 tons (2.3 Mg) of dry solids per day required.

## Alternate Units for Consideration or Evaluation

Any of the twenty or so varieties of continuous BFP's available from 11 different companies. Depending on the length of the truck haul and the cake dryness requirements for final disposal the design engineer would prescreen the many alternates and select perhaps three companies to work with in proving specific devices and carrying out bench and pilot scale qualification trials.

For the purposes of this example it will be assumed that a dry solids content cake of at least 28 percent is required. Accordingly, units such as the R. B. Carter Series 31, Komline Sanderson Unimat, Parkson Magnum Press, Ashbrook-Simon Hartley Winklepress, and Carborundum Sludge Belt Filter Press would certainly be considered. Certain models of the Tait Andritz, Infilco Degremont Floc-Press and Passavant Vac-U-Press would require at least preliminary consideration with further study dependent on estimates of capabilities from the supplier firms.

## Evaluation Procedure

The systematic procedure for evaluation would be identical to that described in the RVF design example.

## Bench Scale Tests

Most of the equipment suppliers have laboratory or bench scale test equipment and procedures which indicate general acceptability of their units. In most cases, unless the sludge to be dewatered is an unusually easy one, pilot scale testing will yield much more accurate design criteria and should be pursued. Most companies have mobile pilot or full size units.

## Design Calculations

1. Operating cycle to be 35 hours per week (7 hours/day), permitting start-up and wash down times within 8 hour shift.
2. One CBFP with adequate spare parts to be maintained.
3. Size of CBFP.—Production rate proves to be 50 GPM (3.2 l/s) of 3–4 percent feed sludge giving rate of 228–305 lb/hr/ft (341–455 kg/hr/m) width (from pilot test runs). Solids capture is an acceptable 93–98 percent in all tests. Cake solids with complete press (all sections, including high pressure stage) in use is 38 percent. Without high pressure section, cake solids are 30 percent. Polymer dosage is consistent. Design Engineer must then assess added capital and O/M costs for high pressure section and effect of 8 percent drier cake on haulage costs to determine which unit is to be chosen. A single CBFP of two meter width would be adequate if several days sludge storage surge capacity were provided. Alternatively 2 one meter wide units could be chosen.
4. Sizing of auxiliary equipment.—Same as described in RVF design example. If, for example, a Komline Sanderson Unimat were the selected unit, the basic machine is just under 24 feet (7.3 m) long, width requirement is 5 feet 2 inches (1.6 m) at base with the upper drive motor making upper width need just under 8 feet (2.4 m). Height of the Unimat is 10 feet 2 inches (3.1 m). The same considerations apply to selection of a suitable flocculant system, sizing of conditioning system and overall "Dewatering System Considerations" as noted in the RVF design example.

# DESIGN EXAMPLE—CONTINUOUS BELT FILTER PRESS—40 MGAL/D (1.75 $m^3/s$) PLANT

## Basic Assumptions

1. Anaerobically digested mixture of primary and W.A.S. at 4 percent dry solids content, 60 percent primary and 40 percent W.A.S.
2. Ultimate disposal by either composting or incineration, both systems requiring a minimum cake solids content of 30 percent.
3. The sludge removal rate to be an average of 25 dry tons (22.7 Mg) of solids per day.

## Alternate Units for Consideration

Same comment as in 4 Mgal/d (0.18 $m^3/s$) example preceding.

## Evaluation Procedure

The same procedure as described in the RVF design example could be used, except:

1. Determination of the calorific value of the dewatered cake produced in pilot tests would be essential for evaluating efficacy of incineration and to ensure whether or not autogenous incineration would be achieved in burning periods (there is no such thing as totally autogenous incineration since startup and shutdown procedures require fuel usage regardless of cake characteristics). Nonetheless, self-sustaining combustion would at least minimize fuel consumption.
2. Review of the suitability for composting could be carried out with experts in that field.

## Bench Scale and Pilot Tests

Same as in 4 Mgal/d (0.18 $m^3/s$) example.

## Design Calculations

1. Pilot results show that 50 gal/min (3.2 l/s) of 3–4 percent sludge will yield a cake solids of 38 percent at a production rate of 228–305 lb/hr/ft (341–455 kg/hr/m) width, with adequate 93–98 percent solids capture and usage of polymer at $10 per ton ($11/Mg) of dry solids.
2. Operating cycle.—To be based on 3 shifts/day, 7 days per week and 22 hours/day unit operating time since incineration requires continuous operation to minimize fuel consumption.
3. Sizing of CBFP.—50,000 pounds/day (22,730 kg/day).

| *Meter width* | *Daily production/unit (pounds)* |
|---|---|
| 1 | 16,500 |
| 2 | 33,000 |
| 3 | 49,500 |

   On the above basis 4 one meter units or 2 two meter units would be chosen.
4. Summation.—All other facets of the design procedure would be similar to the 4 Mgal/d (0.18 $m^3$/s) RVF design example.

## PRESSURE FILTERS

The original main focal point for the development of the plate and frame, and recessed chamber types of pressure filters was Stoke-on-Trent, United Kingdom. The slurriers incident to the manufacture of pottery and china are particularly difficult to dewater and as a result pressure filters were employed.

These types of pressure filters, particularly the recessed chamber type have been frequently designed into the U.K. wastewater treatment plant sludge dewatering systems.

A few U.S. installations of pressure filters have also been made in the past few years.

Pressure filters are batch devices and to some extent because of the level of development of feed and chemical dosage systems normally use substantial quantities of metal salt and lime for conditioning. These chemicals require relatively extensive handling systems requiring considerable maintenance. This is one of the factors which has slowed acceptance of pressure filters outside the United Kingdom.

Essentially, a pressure filter consists of a series of vertical plates, usually recessed, covered with cloths to support and contain the cake, mounted in a framework consisting of head supports connected by two heavy horizontal and parallel bars or an overhead rail. Figure 6–23 shows a cross section of a pressure filter.

Conditioned sludge is pumped into the pressure filter at increasing pressure. Presses are normally supplied to operate at either a nominal 100 lb/in.$^2$g (7 kg/cm$^2$) or 225 lb/in.$^2$g (16 kg/cm$^2$). Cake building time or sludge feed time is normally 20 to 30 minutes followed by a 1 to 4 hour pressing period. The press is then opened and the filter cake falls off into the removal system.

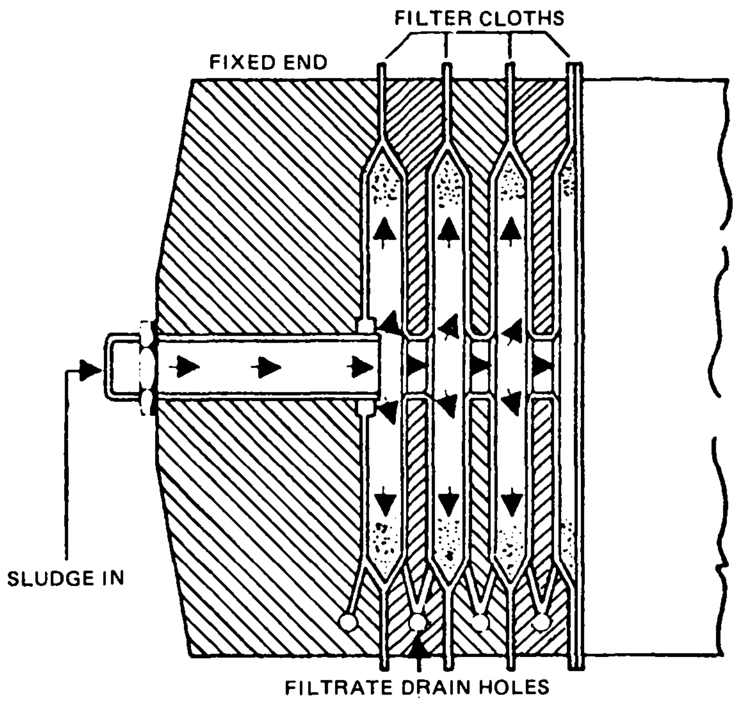


Figure 6–23.—Cross section of a partial pressure filter.

While pressure filters will generally produce a cake solids content 10–20 percent points drier than a rotary vacuum filter, some portion of these total cake solids are lime and metal salt rather than sewage solids. Capacities of pressure filters are usually about 10 to 20 percent of the loadings achieved on rotary vacuum filters.

Significant developments in Pressure Filter technology are the diaphragm press and other membrane type presses which are discussed later.

Since an excellent survey of three operating U.S. installations was available, a review of those case histories is the most applicable way to present a perspective on conventional recessed chamber type presses.

## CASE HISTORY—KENOSHA, WISCONSIN

This is a 26 Mgal/d (1.1 $m^3$/s) plant with a primary and activated sludge system.

1. The sludges are mixed, gravity thickened, anaerobically digested, and then dewatered in Nichols (Edwards & Jones) pressure filters. The dewatered cake is given to farmers who land spread from manure spreaders.
2. Chemical dosage is 3 percent ferric chloride and 25 percent lime (both on a dry solids sludge basis).
3. Digested sludge at 3–7 percent solids is dosed in line with ferric chloride and lime is added in a subsequent mix tank with slow speed mixing.
4. Two Moyno pumps feed the two presses simultaneously. The Moynos have worked very well. Filtrate is returned to head of plant.

5. Cycle includes maintenance of 100 lb/in.$^2$g (7 kg/cm$^2$) for 30 minutes and total cycle time is 2-1/3–2-1/2 hours. Operate 16 hours per day, 7 days per week to produce 12 tons (10.9 Mg) per day of dry solids cake at 35–38 percent solids. Cake thickness is one inch.
6. Two Nichols-Edwards & Jones pressure filters, with 80—4 feet by 4 feet (1.2 m × 1.2 m) plates (rubber-coated steel) used.
7. One operator in continuous attendance.

## Results

Table 6–30 shows the good handleable press cake and clear filtrate.

## Problems

High chemical dosage and costs have been experienced. Cake is actually about 25 percent added chemical so analysis is really about 65 percent water, 26 percent sewage sludge and 9 percent inorganic chemical. Net sludge production must be reduced by 25 percent to get actual figures. Excessive wear in cloths and stay bosses causes serious maintenance problems. Filter cloths were replaced 3 times in 2 years ($3,000 per press per change). Severe ammonia odor problems have occurred in press room (effect of lime and high pH).

## Comment

Despite problems noted above there have no extensive forced downtime periods in the 2 years of operation. Much of the chemical consumption might be eliminated if the alkalinity of the digested sludge were washed out in a properly designed and operated elutriation system using flocculants. Why use pressure filters when the wet cake is disposed of on land by a manure spreader?

## BROOKFIELD, WIS.

This plant design includes a primary and activated sludge system and contact stabilization. Flow is 2 Mgal/d (0.09 m$^3$/s). 80 percent Primary Sludge + 20 percent Secondary Sludge is mixed, pumped through a grinder, diluted with recycled incinerator ash (0.5 lb/lb sludge), conditioned with lime (15–18 percent) and Ferric Chloride (5–7 percent), pressed and fed to a 5 hearth incinerator. 95 percent of incinerator ash is recycled. The incineration is not autothermic and uses natural gas. Pressure filters are standard Passavant design with forty-six 52″ (1.3 m) diameter plates of steel and have been operated for 1-1/2 years.

Table 6–30.—Costs—pressure filtration, Kenosha, Wis.

| Costs | $/ton |
|---|---|
| Labor | $7.43 |
| Chemicals | 20.17 |
| Power | 1.71 |
| Maintenance | 3.25 |
| Total | 32.56 |

## Results

Plant personnel state that no major operating problems have been encountered. There have only been two "Sludge Blowing Incidents" in the 1-1/2 years of operation. Press cloths have had to be replaced every 6 months at a cost of $3,600 per shot. The press cake, which contains a large amount of inorganic conditioning agents and recycled ash averages 45 percent *total* solids. The press cake is only 30–40 percent volatile so the ratio of water/sewage solids is quite high.

## Comments

1. The mixed sludge being processed is a relatively easily dewaterable material which is high (80 percent) in primary content and high in fibrous material. Indeed the high fiber content has caused problems in the press cake breaking operation.
2. No records are available on natural gas consumption and no cost data on the system have been made available.
3. The system appears to be a complex high capital and high operating and maintenance cost one which is difficult to rationalize, particularly at a plant with such an easily processable sludge.
4. The plant has two components of interest to other potential press filter designs: the wet sludge grinder and the slow speed cake breaker.

## Conclusions on U.S. Results to Date

Reference 16 from which the above results came, is an excellent review of the current U.S. installations.

The conclusions from reference 9 are as follows:

1. In looking at the two types of presses, we found some advantages with the lower pressure design. Essentially, it is a much simpler operation. The recycling of incinerator ash seemed to provide few benefits, particularly because it only complicated the operation with additional material handling equipment.
2. In general, we found that filter presses are an acceptable method for dewatering sludge. Theoretically, they should always produce an autocombustible sludge cake. But, practically, we know of no installation anywhere that can achieve this. The ash recirculation is probably the limiting factor. (The inorganic conditioning agents also contribute to the problem.)
3. Filter presses seem to be quite capable of handling different sludge concentrations and different types of sludge feed. Proper conditioning, especially with lime, is the key to good operation. Vacuum filters are not quite so adaptable.

4. The necessity of using high lime for conditioning could be a drawback. Lime handling is always difficult.
5. Prior to a large scale installation, pilot plant work should always be performed to evaluate the dewatering characteristics and chemical requirements.
6. Filter presses have a higher capital cost than vacuum filters. The presses also usually have a higher operational cost. Their real advantage is in greatly reducing the costs of final disposal for the sludge cakes. A detailed economic analysis of the total system is needed before deciding for or against filter presses.

## POLYELECTROLYTE CONDITIONING FOR PRESSURE FILTERS

Due to the more prevalent previous incidence of the use of filter presses in continental Europe and the United Kingdom, and also due to innovative work there, the successful use of certain polyelectrolytes in conditioning sludges for dewatering in pressure filters has been realized at a number of locations.

## Farnham Pollution Control Works, Thames Water Authority, U.K.

This plant is a primary and trickling filter installation. Humus sludge is recirculated to the primaries, the mixed sludge gravity thickened, and then dewatered on two filter presses. Operating pressures are 85–100 lb/in.$^2$g (6–7 kg/cm$^2$).

Initially the plant used aluminum chlorohydrate for sludge conditioning. Figure 6–24 is a flow diagram of the dewatering system.

The Farnham plant experienced severe filter cloth blinding problems and proceeded to carry out diagnostic trails with various conditioning agents to rectify the problem. They found that by converting the system to use Allied Colliods Zetag 63 polyelectrolyte the cloth blinding problems were alleviated sufficiently for the two presses to cope with the sludge load. (See Table 6–31 for dewatering results.)

## CASE HISTORY—THORNBURY STP, U.K.

Reference 17 describes exhaustive test work on the use of polymers for conditioning sludge for dewatering via recessed chamber pressure filters.

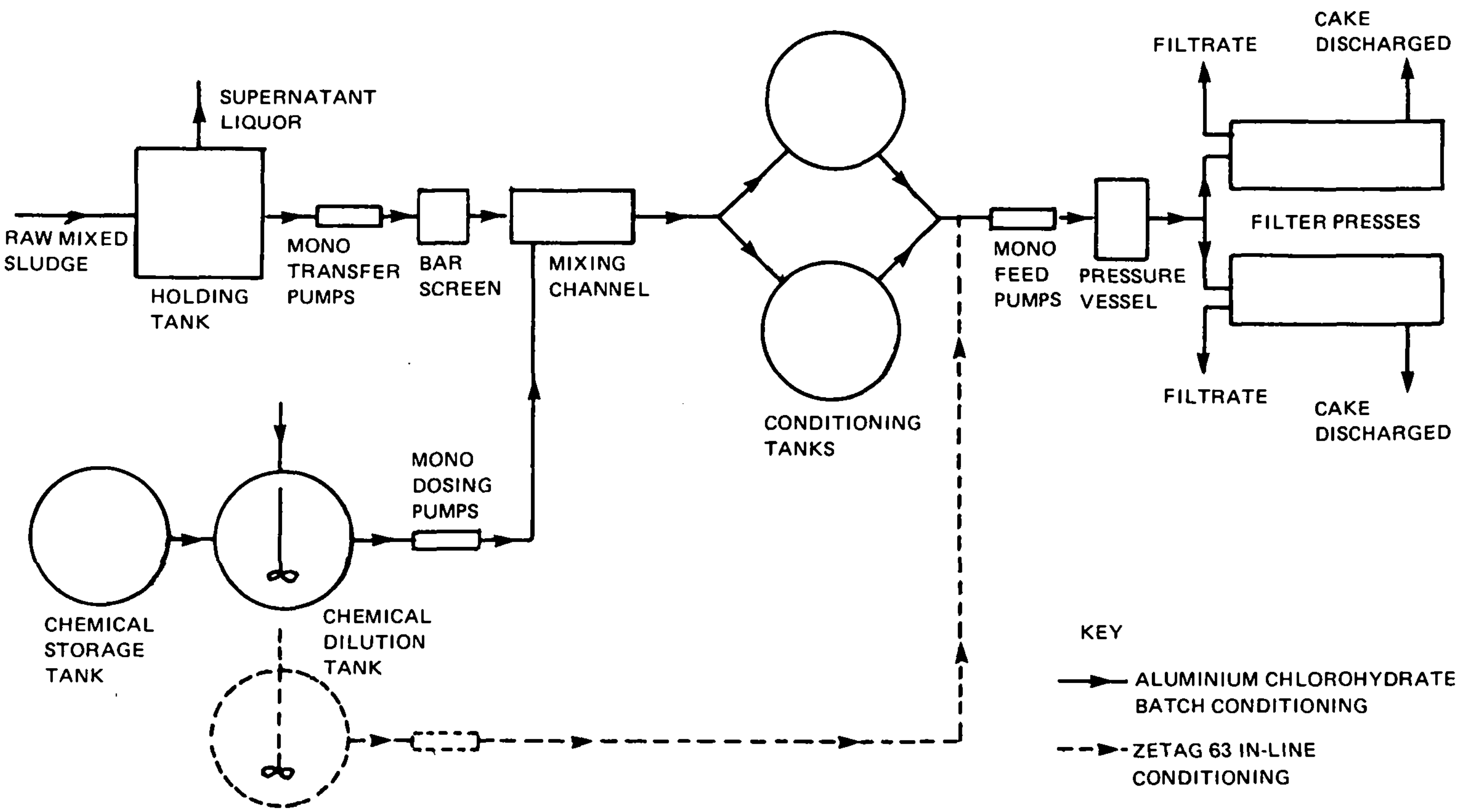


Figure 6–24.—Farnham Plant dewatering system.

Table 6–31.—Farnham dewatering results

| Conditioning agent | Dose (% on dry solids) | Cost ($/ton dry solids) | CST range during cycle (seconds) | Pressing cycle time range (hours) |
|---|---|---|---|---|
| Aluminum chlorohydrate (batch)... | 2.5 | 22.00 | 10–65 | 6–18 |
| Aluminum chlorohydrate (in-line)... | 2.5 | 22.00 | [a]— | 6–12 |
| Zetag 63 (batch)................... | 0.2–0.3 | 6.70–10.10 | 10–32 | 6–9 |
| Zetag 63 (in-line)................... | 0.2–0.3 | 6.70–10.10 | 8–14 | 3–6 |
| Ferric chloride and lime (batch)... | 3<br>25 | 14.80 | 8–45 | 3–13 |
| Ferric chloride and lime (in-line)... | 3<br>25 | 14.80 | 8–15 | 3–5 |

[a]Results not available.

By virtue of using in-line conditioning and observing logical procedures the results shown in table 6–32 were achieved.

The Thornbury works processes a mixture of 45 percent primary sludge and 55 percent of mixed sludges from adjacent secondary treatment plants. In addition to illustrating successful use of polyelectrolytes, the article delineates other significant facts relative to pressure filter design.

## MEMBRANE USE—PRESSURE FILTERS

References 18 and 19 describe the successful upgrading of the production rate in conventional recessed chamber pressure filters by equipping same with alternate "membrane" plates. This retrofitting process causes each of the chambers formed between the standard recessed plate and the membrane plate to be subject to the squeezing action of a membrane at will during the press cycle. The membrane plate is a steel reinforced rubber plate in which the rubber membrane is inflatable by air pressure. After the initial filling period in a press cycle, when the filtrate rate falls off, the sludge feed pump is stopped and the membrane inflated to give a pressure up to 150 lb/in.$^2$ (10.5 kg/cm$^2$) to squeeze the partially formed cake and obtain quick dewatering.

Table 6–32.—Thornbury, U.K.—pressure filtration

| Conditioner | Percent dry solids | | Conditioner cost ($/ton dry solids) | Press cycle (hrs.) |
|---|---|---|---|---|
| | Feed | Cake | | |
| Aluminum chlorohydrate............ | 4.6 | 38 | 23.40 | 4.9 |
| Polyelectrolyte (Zetag 94)......... | 4.6 | 37 | 4.60 | 4.9 |
| (Primary + secondary sludge) | | | | |

Table 6–33.—Conventional versus membrane press Severn Trent Water Authority[18]

| Type press used | Cycle (minutes) | Cake thickness (inches) | Dry solids (percent) | Weight (lbs) | Output (lb/hr/press) |
|---|---|---|---|---|---|
| Conventional.... | 390 | 1.25 | 28 | 1,227 | 186 |
| Membrane ...... | 87 | 0.7 | 27 | 558 | 385 |
| (Raw feed sludge—3.9% dry solids—2.0% alum chlorohydrate cond.) | | | | | |

As can be seen in table 6–33 following, though a thinner cake results, the overall filtration cycle is so much shorter that the total throughput doubles or even triples in some cases.

The suppliers of the rubber membrane plates claim that new installations of the membrane type unit are less expensive overall due to the increased capacity of the membrane units.

A somewhat analogous but different type of variable volume pressure filter is described in the following section.

## DIAPHRAGM TYPE PRESSURE FILTERS

As described in reference 20, a new type of pressure filter, employing flexible rubber diaphragms between the chambers of a pressure filter, has recently been introduced into the United States. This type device was developed in Japan and there are several operating installations there.

At least two versions of this new type of pressure filter have been tested and are available in the United

States. The earliest one was supplied by NGK Insulators Ltd., of Nagoya who have now licensed Envirex division of Rexnord for U.S. sale of their device. Ingersoll Rand has the U.S. rights to the Lasta automatic diaphragm pressure filter. There are indications that Dart Industries and Industrial Filters OMD of Chicago have devices based on similar principles.

Figure 6–25 is a diagram of the I. R. Lasta press that illustrates the operating principles.

As will be noted in figure 6–25 the feed slurry enters the top of the chamber between the filter cloths and gradually fills the chamber. After a cake is formed the diaphragm is expanded by water under pressure to 250 lb/in.$^2$g (17.6 kg/cm$^2$) which squeezes and dewaters the cake. The filter plates are then automatically opened and the cake discharged. Cloth washing ensues before another pressing cycle.

It is claimed that the length of the cycle is shorter than for conventional presses because of the improved control of the relationship between cake formation and pressure build-up.

Table 6–34 lists dimensional data on the I. R. Lasta press.

The most detailed report on these devices is Reference 20 which describes the extensive pilot work at Blue Plains with the Envirex-NGK Locke diaphragm press. This Envirex unit is highly automated and in work at Blue Plains (mixture of primary and W.A.S. sludges), it produced a 40 percent total dry solids cake using 20 percent lime and 10 percent ferric chloride dosages. The only problem is that when the correction is made for the inorganic conditioning solids present in the dewatered cake, the percentage of dry sewage solids in the cake relative to water content is only about 28 percent.

This new type pressure filter does offer much improved capabilities over conventional pressure filters for extremely difficult to dewater sludges. Pricing figures available indicate that the units will be priced about eight times the price of a conventional pressure filter, so the need must be clear and obvious.

## CENTRIFUGES FOR DEWATERING

Horizontal solid bowl decanter type centrifuges have been used for wastewater sludge dewatering for a number of years. They were popular for primary sludges with low grit content in coastal resort areas with large swings in loadings because of ease of operation, quick startup and shutdown and ease of odor control. Attempts to adapt these relatively high speed devices (g forces of 1000+) to heavy duty operation in large cities or for use with mixed sludges containing significant quantities

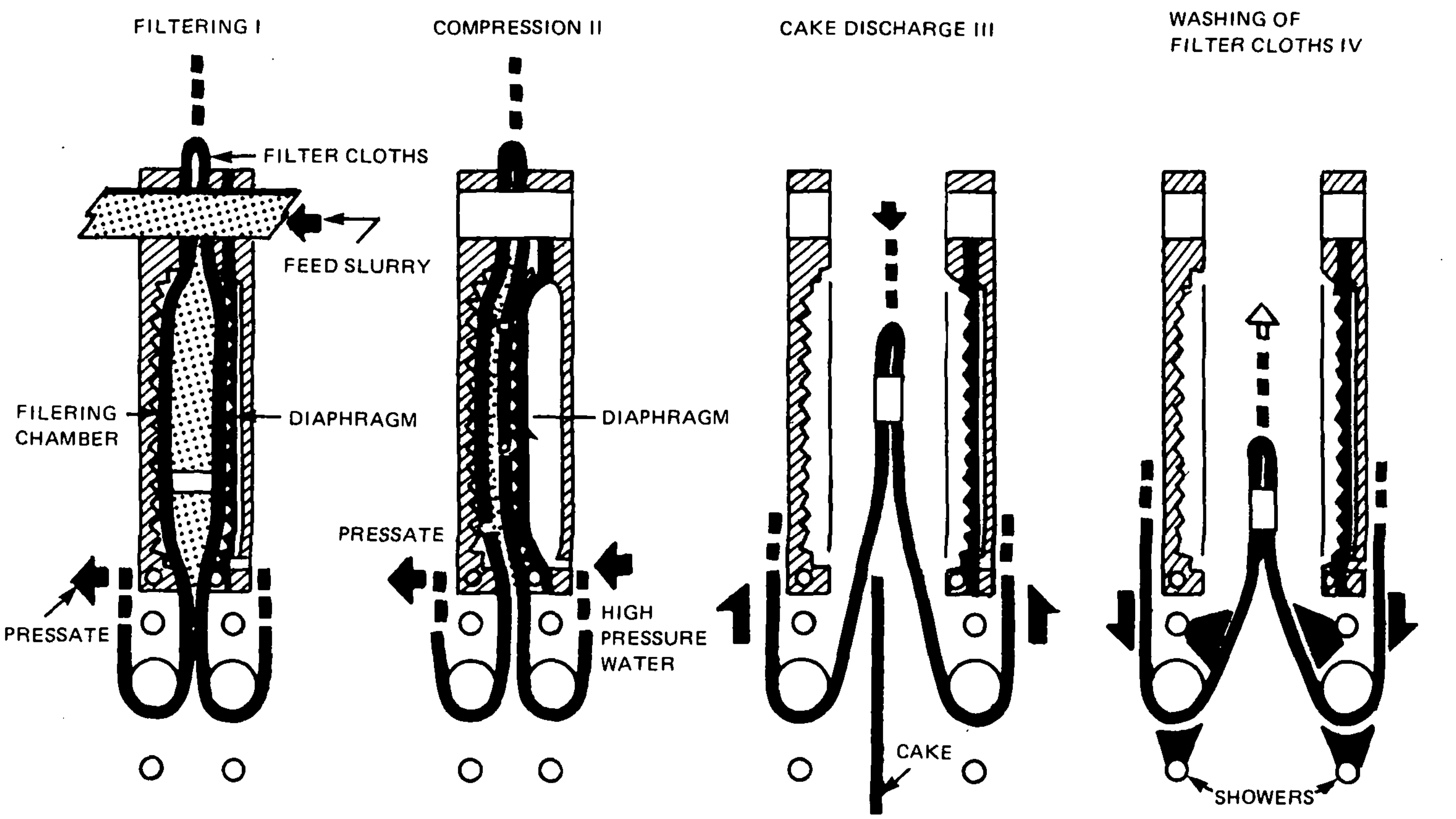


Figure 6–25.—I. R. Lasta diaphragm pressure filter.

Table 6–34.—I. R. Lasta automatic filtering press

| Size of filtering plate | Number of filtering chambers | Filtering area | | Height | | Length | | Width | |
|---|---|---|---|---|---|---|---|---|---|
| | | m² | ft² | mm | ft. | mm | ft. | mm | ft. |
| 600 mm (24")......... | 8 | 4 | 43 | 2,050 | 7 | 2,660 | 9 | 1,610 | 5 |
| | 14 | 7 | 75 | | | 3,650 | 12 | | |
| | 20 | 10 | 108 | | | 4,640 | 15 | | |
| 800 mm (32")......... | 14 | 13 | 140 | 2,485 | 8 | 3,490 | 13 | 1,800 | 8 |
| | 20 | 19 | 204 | | | 4,930 | 16 | | |
| | 26 | 24 | 258 | | | 5,920 | 19 | | |
| 1,000 mm (40")........ | 20 | 30 | 323 | 2,845 | 9 | 5,240 | 17 | 2,100 | 7 |
| | 26 | 39 | 420 | | | 6,230 | 20 | | |
| | 32 | 48 | 516 | | | 7,220 | 24 | | |
| 1,250 mm (50")........ | 26 | 62 | 667 | 3,200 | 10 | 6,555 | 22 | 2,600 | 9 |
| | 32 | 77 | 829 | | | 7,545 | 25 | | |
| | 38 | 91 | 979 | | | 8,535 | 28 | | |
| 1,500 mm (60")........ | 32 | 112 | 1,205 | 3,620 | 12 | 8,205 | 27 | 3,050 | 10 |
| | 38 | 113 | 1,431 | | | 9,225 | 30 | | |
| | 44 | 154 | 1,657 | | | 10,245 | 34 | | |

of biomass were previously plagued by two problems:

1. Erosion of the surfaces exposed to high speed impingement of abrasive materials caused maintenance problems.
2. Prior to the development of polyelectrolytes capable of providing a reasonable clean centrate and avoiding serious fines recirculation problems, solids capture was inadequate.

In the past 5 years or so six steps were taken which have helped this type device gain a wider use:

1. Development and use of new high molecular weight cationic shear resistant polyelectrolytes.
2. Use of lower rotational speeds to reduce turbulence, power costs, and erosion wear problems.
3. Use of a concurrent flow pattern for sludge and centrate to minimize turbulence.
4. Adjustable variation of speed differential between the bowl and the sludge removal scroll.
5. Use of longer bowls with smaller diameters.
6. Provision of extremely large units at plants with large sludge removal needs producing an economy of scale.

Various manufacturers have combined some of the above features in their newer models. This resulted in a surge of popularity about 4 years ago. Since the energy crisis the degree of popularity of centrifuges, even with the above mentioned improvements, has slackened because of energy costs.

Once again, the pioneering development work on these devices was carried out primarily in West Germany. The most practical description of these developments is contained in references 21 and 22 which are excerpted in the following section.

## CASE HISTORY—CENTRIFUGAL DEWATERING—WUPPERTAL-BUCHENHOFEN, GERMANY

Reference 21 is a comprehensive article relating results obtained at Wuppertal-Buchenhofen plant with a low speed concurrent flow type unit. This is a combined municipal-industrial treatment plant treating 1,200,000 population equivalent. After primary and biological treatment the mixed sludges are thickened to 3–4 percent and anaerobically digested, followed by sludge settlement and decantation, thence dewatering.

After initial trial work the authority asked for competitive tenders from various suppliers of centrifuges with performance requirements as follows:

1. Capacity of each centrifuge: 40–60 $m^3$/hour of sludge with feed of 2.5–3 percent dry solids.
2. Minimum cake solids: 20 percent.
3. Centrate maximum suspended solids of 0.2 percent.
4. Maximum polyelectrolyte dosage permissible of 3.6 kg/Mg of dry solids (100 gm/$m^3$).
5. Maximum permissible power consumption of 1 kWh per cubic meter of sludge feed including ancillary equipment such as pumps, flocculant metering stations, etc.
6. Guaranteed life of screw conveyor = 10,000 hours.
7. Provision of a package plant with a minimum capacity of 40 $m^3$/h for a 4-month trial period under a leasing agreement.

Table 6–35.—Effect of speed differential on throughput and dry solids

| Speed differential | 2 | | 4 | | 6 | |
|---|---|---|---|---|---|---|
| Flocculant dosage (g/$m^3$) | 60 | 80 | 60 | 80 | 60 | 80 |
| Dry solids carried by discharge, percent | 26 | 28.5 | 24 | 23 | 20.5 | 20 |
| Dry solids carried by centrate (undissolved solids) | 0.35 | 0.25 | 0.17 | 0.07 | 0.12 | 0.07 |
| Ideal throughput ($m^3$/h) | 33 | 37 | 43 | 45 | 40 | 48 |

KHD Industrieanlagen AG Humboldt-Wedag of Cologne (U.S. Licensee—Bird Machine) won the contract and initially installed two S3-2 type low speed concurrent flow centrifuges with capacities of 20–30 $m^3$/h each. These units met the agreed performance guarantees but when the full civil installation was completed they were replaced, as planned, by two of the larger S4-1 units (of the same basic type) but with capacities of 40–60 $m^3$/h each.

Power consumption for the complete dewatering plant was 0.9–0.95 kWh/$m^3$ with S3-2 units and improved to 0.75–0.8 with the larger S4-1 units. Disage of Zetag 92 polymer (Allied Colloids) averaged 60–80 gm/$m^3$.

The article contains much data on the effect of centrifuge dewatering variations on overall process performance and sludge disposal costs.

A significant factor studied was that of the effect of the differential in speed between the scroll and the bowl (see table 6–35).

As can be seen in table 6–35, a 28.5 percent dewatered cake at a reasonable throughput of 37 $m^3$/hour and centrate suspended solids of 0.25 percent can be obtained with flocculant dosage of 80 g/$m^3$ by using a speed differential of 2 instead of 6.

The paper claims and purports to show that very large capacity centrifuges of the improved low speed-concurrent flow type, when operated in a lower differential speed mode can offer significant capital and O/M cost savings where large volumes of sludge are to be processed.

Unit costs are given as follows:

Operating—Deutsche mark 36.40/ton (DM 40.12/Mg) dry solids
Annual Capital—Deutsche mark 47.60/ton (DM 52.47/Mg) dry solids

## CASE HISTORY—CENTRIFUGAL DEWATERING—STOCKHOLM, SWEDEN

Stockholm has operated three high speed centrifuges for a 3 year period and also has operated a new low speed concurrent flow unit on the same sludge for 1-1/2 years.

Table 6–36 shows the results obtained with the two different types of centrifuge.

Table 6–36.—Side by side comparison—process results

| Centrifuge design sludge identification | Anaerobically digested primary plus waste activated with alum sludge: Low speed | High speed |
|---|---|---|
| No. of operation units | 1 | 3 |
| Flow rate per unit | 190 gal/min | 90 gal/min |
| Percent feed consistency | 3 | 3 |
| Percent cake solids | 16–18 | 16–18 |
| Percent solids recovery | 95–98 | 95–98 |
| Polymer type | Allied Colloids Cationic | Percol #728 |
| Polymer dosage | 6 lbs/ton | 12 lbs/ton |

While table 6–36 only shows the improvement realized by reduction in polyelectrolyte costs by about $9/ton ($10/Mg) (which is a considerable savings), table 6–37 illustrates the additional advantages for the low speed design.

Wear played an important part in displacing the high speed centrifuges in favor of the low speed centrifuges at this particular plant. The low speed centrifuge was inspected after 2,000 hours of operation and found to have only 1/18 of the wear of the high speed alternative. The abrasive protection on the low speed machine conveyor blades is tungsten carbide, while the protection on the high speed machine is equivalent to an alloy called Stellite 1016. The Stellite material is considered inferior to the tungsten carbide hardness values approach Rc-69. Experience shows that if both materials had been similar that the wear rate would still have favored the low speed design by as much as a five to one ratio.

Summarized in table 6–38 is the annual cost analysis of the operation of these two types of centrifuges installed side by side. The low speed unit clearly has the

Table 6–37.—Side by side comparison machine parameters

| Centrifuge design | Low speed | High speed |
|---|---|---|
| Bowl diameter | 36″ | 25″ |
| Bowl length | 96″ | 90″ |
| Centrifugal force | 511 × G | 1,878 × G |
| Unit flow rate | 190 gal/min | 90 gal/min |
| Unit pool volume | 196 gallons | 73 gallons |
| Sigma factor | $1.15 \times 10^7$ $cm^2$ | $5.3 \times 10^7$ $cm^2$ |
| Unit motor size rating | 100 hp | 180 hp |
| Absorbed horsepower | 0.3 gal/min | 0.6 gal/min |
| Noise level at 3 ft | 80–85 dBa | 95–100 dBa |
| Wear at 2,000-hour inspection | 1/2 mm | 9 mm |

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Table 6–38.—Side by side comparison annual cost—profile

| Centrifuge design | Low speed | High speed |
|---|---|---|
| Tons/year per unit | 12,483 | 5.913 |
| Power expenditure | $0.06/ton | $1.19/ton |
| Polymer expenditure | $9.00/ton | $16.00/ton |
| Maintenance expenditure | $1.21/ton | $8.30/ton |
| Amortized equipment | $1.50/ton | $2.44/ton |
| Total annual cost | $12.33/ton | $27.93/ton |

Table 6–39.—Dimensional data—low speed centrifuge

| Model No. | Overall length (in.) | Overall width (in.) | Overall height (in.) | Weight (lbs) |
|---|---|---|---|---|
| HB 2500 | 138 | 80 | 36 | 6,500 |
| HB 3700 | 139 | 72 | 41 | 9,400 |
| HB 6400 | 276 | 150 | 71 | 3,440 |

edge in all categories. Power consumptions are one-half (1/2) that of high speed unit. With respect to polymer consumption, the low speed centrifuge in this particular case utilized 44 percent less cationic polymer than the high speed centrifuge. With respect to conveyor maintenance, we have modified the high speed centrifuge figure to reflect a ratio of conveyor resurfacings more in the category of 5 to 1 than the 18 to one margin indicated by the actual side by side installation. The category entitled "Amortized Equipment" includes the cost of centrifuge, the motor, and the starter, and is expressed on a tonnage basis and reflects an amortization rate of 7 percent interest over a 20-year period. Electrical usage rate was assumed to be 0.02/kWh and polymer (Allied Colloids Percol 728) was figured at $1.50/lb ($3.30/kg).

While the larger size of the low speed unit would account for a minor portion of the above noted superiority, it is abundantly clear that the lower speed concurrent flow unit is superior from a cost-effectiveness standpoint.

## Dimensional Data—Centrifuges

Table 6–39 shows dimensional data for one brand of the newer low speed centrifugals.

## CASE HISTORY—CENTRIFUGAL DEWATERING—BURLINGTON, WIS., WWTP

The experiences at Burlington are described (in an outstanding fashion) in reference 23.

The Burlington plant treats an average flow of combined municipal and industrial wastes at DWF level of 1.5 Mgal/d (0.06 $m^3$/s) and a wet weather flow of 2.0 Mgal/d (0.09 $m^3$/s).

The treatment plant employs contact stabilization (12 hour aeration time, 25 percent return rate, MLSS of 2000 mg/l). The F/M ratio is 0.2 to 0.5. A sludge age varying from 5 to 12 days is employed, including aeration and aerobic digestion time.

The above described liquid treatment system results in sludge disposal requirements of 160,000 gallons (606 $m^3$) of W.A.S. per week or 3400 pounds (1545 kg) per day, about 27,000 gallons/day (102 $m^3$/day).

The plant was designed for ultimate liquid sludge disposal by lagoon. When this disposal option was curtailed, dewatering studies ensued. Needless to say, the sludge dewatering problems are significant. It is a classic example of the problems which result when a plant's liquid treatment system is designed for liquid sludge disposal and then dewatering is required.

A batch, cycling, basket centrifuge was tested, purchased, installed and has been operated for some time. The essence of the results of the full scale performance is listed in table 6–40.

Table 6–40.—Basket centrifuge operation—Burlington, Wis., WWTP

| | | |
|---|---|---|
| Feed rate (gal/min) | 23 | 88 |
| Dewater rate (lbs D.S./hr) | 104 | 397 |
| Hours required/week | 168 | 44 |
| Labor + trucking cost ($/wk) | 378 | 99 |
| Electricity cost ($/wk) | 147 | 48 |
| Chemical cost ($/ton) | 0 | 30 |
| Cake solids, percent | 6–8 (U.T.) | 13–15 (T) |
| Skimming volume, percent | 50 | 14 |
| Total costs | 62 | 47 |

As can be seen, despite the high polymer cost, the overall cost analysis showed the total operation to be more cost effective with polymer use.

It should be noted that if the city could start again from square one, it is certain that, now having to dewater sludge, and knowing the overall energy costs of the type total system involved, a different liquid treatment system would be chosen.

Additional valuable insights in the referenced paper relate to the correlations between activated sludge system operating parameters and resulting sludge processability.

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