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Design Manual

Dewatering Municipal Wastewater Sludges

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Notice

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This document is not intended to be a guidance or support document for a specific regulatory program. Guidance documents are available from EPA and must be consulted to address specific regulatory issues.

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Chapter 1
Introduction

1.1 Purpose and Scope

This manual presents up-to-date information on dewatering processes for municipal wastewater sludges. The design engineer can use this document to aid in the selection of an appropriate dewatering process for a particular application. The manual revises and updates the information on sludge conditioning and dewatering previously found in the U.S. Environmental Protection Agency's *Process Design Manual for Sludge Treatment and Disposal* (October 1979) and *Process Design Manual for Dewatering Municipal Wastewater Sludges* (October 1982). Significant advances have been made in dewatering technology since preparation of these documents. Also, the regulatory criteria for disposal of sludges by landfilling, combustion, land application, and ocean disposal have been tightened.

This manual considers the upgrading of existing dewatering processes, as well as the designing of new ones, and pays particular attention to the needs of small facilities. All currently employed technologies are considered as well as emerging ones. Proper sludge management requires the use of several sludge treatment and disposal processes. Selection of a dewatering process is not an independent step, nor is it even the initial step. This selection process is guided by many factors, including the public's desires, the final disposal option selected, regulatory requirements, and the size of the facility. Table 1-1 shows other EPA Technology Transfer publications which may be used to complement this one when preparing the feasibility design of a sludge management system.

Design parameters, performance capabilities, and design deficiencies for all dewatering processes are presented. While some cost information (in the form of a range of values) is included in certain sections of this manual, the intent is not to give detailed cost estimating information. Such information can be found in the Technology Transfer *Handbook for Estimating Sludge Management Costs at Municipal Wastewater Treatment Facilities* (EPA 625/6-85-010). Some specific cost information is, however, presented with the case studies and design examples. The manual specifically discusses those processes where the most extensive and cost-effective dewatering performance improvements have been made, including air drying, centrifugation, belt press filtration, and recessed plate pressure filtration using polymer conditioning.

Table 1-1. EPA Technology Transfer Sludge Management Publications

Process Design Manuals

1. Municipal Sludge Landfills (October 1978)
2. Sludge Treatment and Disposal (October 1979)
3. Land Application of Municipal Sludge (October 1983)

Seminar Publications

1. Composting of Municipal Wastewater Sludges (August 1985)
2. Municipal Wastewater Sludge Combustion Technology (September 1985)

Brochures

1. Environmental Pollution Control Alternatives: Sludge Handling, Dewatering, and Disposal Alternatives for the Metal Finishing Industry (October 1982)

Handbooks

1. Identification/Correction of Typical Design Deficiencies at Municipal Wastewater Treatment Facilities (October 1982)
2. Estimating Sludge Management Costs at Municipal Wastewater Treatment Facilities (October 1985)

Environmental Regulations and Technology Publications

1. Use and Disposal of Municipal Wastewater Sludge (September 1984)

This document is current as of the summer of 1987 and includes detailed case history information on the newer processes and equipment. These include vacuum assisted dewatering beds; solid bowl centrifuges with backdrive capability and optimized bowl design; third generation belt filter presses; and diaphragm filter presses. This manual describes the capabilities of these and other dewatering processes by presenting data from full-scale field testing and operating installations. In most cases, the information presented is for sludges produced during primary and secondary municipal wastewater treatment. Chemical sludges produced during advanced wastewater treatment are given minimal coverage.

In general, the manual has been prepared for use by experienced engineers involved in the design, selection, and specification of dewatering equipment.

The major types of dewatering processes discussed in this manual include:

- Air Drying Processes (Chapter 6)
  Sand Beds
  Freeze Assisted Sand Beds
  Vacuum Assisted Beds
  Wedgewire Beds
  Lagoons
  Paved Beds
  Other Innovative Processes

- Belt Press Filtration (Chapter 7)

- Centrifugation (Chapter 7)
  High-G Machines
  Low-G Machines

- Vacuum Filtration (Chapter 7)

- Pressure Filtration (Chapter 7)
  Fixed Volume
  Variable Volume

All of these processes are in use today, although a process such as vacuum filtration is rarely seen in a new installation. The manual does not provide detailed discussion of mechanical processes which have been installed at only a few facilities or processes which do not have a proven background of performance.

1.2 Objectives of Dewatering

The general objective of dewatering is to remove water, thereby reducing the sludge volume. This produces a sludge which behaves as a solid and not a liquid, and reduces the cost of subsequent treatment and disposal. In most cases, the percent solids content of a dewatered sludge is set by the requirements for subsequent treatment and disposal; this percent solids content is always significantly higher than the percent solids content of a thickened sludge.

1.3 Location of the Dewatering Process

The combination of processes used for solids treatment prior to dewatering, transport, and disposal varies widely from plant to plant. Generally, however, the dewatering process is preceded by one of the following stabilization processes: anaerobic or aerobic digestion; thickening by either gravity, centrifugation, air flotation, or rotating screen type process; and chemical or heat conditioning. In some cases, raw sludge, particularly raw primary sludge, may be dewatered directly, although the handling and the method of ultimate disposal would have to be considered. After the dewatering operation, further stabilization may be provided by composting; volume and organic reduction may be accomplished by incineration; or the dewatered sludge may be ultimately disposed of by transport to either a landfill or a site for landspreading.

1.4 Using this Manual

This manual has been organized to allow users to concentrate on areas of interest as easily as possible. The following brief chapter and appendix descriptions provide an overview of the manual's organization.

Chapter 2 - Preliminary Considerations
Discusses the size of the treatment facility, regulatory concerns, and performance capabilities of various mechanical dewatering processes. Tables are used to show the percent total solids achievable with different mechanical processes.

Chapter 3 - Sludge Characteristics and Preparatory Treatment
Presents brief discussion of sludge characteristics. Sludges from primary, biological, and chemical wastewater treatment are included. Sludge quantity and quality data are related to how the sludge is produced. In addition, sludge treatment prior to dewatering (such as by thickening and stabilization processes) is discussed in relation to dewatering.

Chapter 4 - Process Selection
Provides guidance for identifying the most cost-effective system from the alternatives presented in Chapters 5, 6, and 7. Considered are the method of disposal, plant size, practicality, and costs. Emphasis is on upgrading and retrofitting.

Chapter 5 - Conditioning
Presents the purpose and methods of sludge conditioning, how the methods work, and how one is selected. Inorganic chemicals, organic chemicals, and thermal conditioning (alone and in combination) are discussed. This chapter also includes tests for selection of chemical dose and determination of dewaterability by different devices.

Chapter 6 - Air Drying Processes
Proven, cost-effective technologies are emphasized. Dewatering processes discussed include sand beds, freeze assisted sand beds, paved beds, vacuum assisted beds, wedgewire beds, sludge lagoons, and emerging systems. For each process, the following topics are discussed: performance data, advantages and disadvantages, problems experienced, design criteria, and O&M concerns.

Chapter 7 - Mechanical Dewatering Processes
Systems are discussed with a focus on theory of operation. This chapter describes belt filter presses,

centrifuges, filter presses, vacuum filters, and emerging processes. Topics discussed include performance capabilities, performance experience, design criteria, and O&M concerns. Wherever possible this chapter provides comparative performance information.

Chapter 8 - Case Studies: Air Drying Systems
Detailed case studies for some of the systems discussed in Chapter 6 are given. Presentations include actual costs for capital equipment, O&M, energy, labor, chemicals, and replacement parts.

Chapter 9 - Case Studies: Mechanical Dewatering Systems
Detailed case studies for some of the systems discussed in Chapter 7 are given. These include high-speed centrifuge operation, low-speed centrifuge operation, recessed plate filter press operation, and belt filter press operation. Presentations include actual costs for capital equipment, O&M, energy, labor, chemicals, and replacement parts. Included are comparative studies of mechanical dewatering devices that were made on either a plant scale or large pilot-plant scale.

Appendix A - Design Examples/Cost Analyses
Examples are presented for upgraded sand drying beds, vacuum assisted drying beds, belt filter presses, solid-bowl centrifuges, and recessed plate filter presses. Step-by-step procedures for design and cost projections are given. Cost information is based on the recently issued EPA handbook on estimating sludge management costs.

Appendix B - Operation and Maintenance: Mechanical Dewatering Systems
This appendix discusses energy requirements; O&M record-keeping (data, frequency, etc.); and simple control tests and observations. The designer can use this appendix when preparing the dewatering system's O&M manual.

Appendix C - Manufacturers and Sources of Equipment
Manufacturers of belt filter presses, centrifuges, and filter presses appear.

Chapter 2
Preliminary Considerations

2.1 Introduction

The quantity of sludge produced in U.S. municipal wastewater treatment plants was last estimated from data obtained in the 1982 EPA Needs Survey. Table 2-1 presents this data for all sizes of treatment plants. The number of Publicly Owned Treatment Works (POTWs) in a particular size category is also given.

It is interesting to note that the smallest plants, <2.5 mgd (0.11 m^3/s), represent 91 percent of the POTWs and produce less than 17 percent of the sludge. In contrast, the largest plants, >100 mgd (4.38 m^3/s) represent less than 0.3 percent of the facilities and produce more than 34 percent of the sludge.

Table 2-1. Estimated Municipal Sludge Production by POTW Size

POTW Size	No. of POTWs	Sludge Produced	Total
mgd		dry tons/yr	percent
<2.5	14,168	1,189,810	17
2.5 - 5	631	515,504	8
5 - 10	352	588,445	9
10 - 20	187	622,478	9
20 - 50	125	924,896	13
50 - 100	40	676,091	10
>100	41	2,324,274	34

ton x 0.9072 = Mg.
mgd x 0.0438 = m^3/s.

A representative survey of U.S. facilities was performed in 1980 by EPA's Office of Solid Waste to determine the choice of sludge use/disposal options by plant size. The results are shown in Table 2-2. The "Other" category frequently means a lagoon or temporary storage facility. Note that small to medium sized facilities more frequently select some form of land use/disposal option than do the large sized facilities, which more frequently use incineration.

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

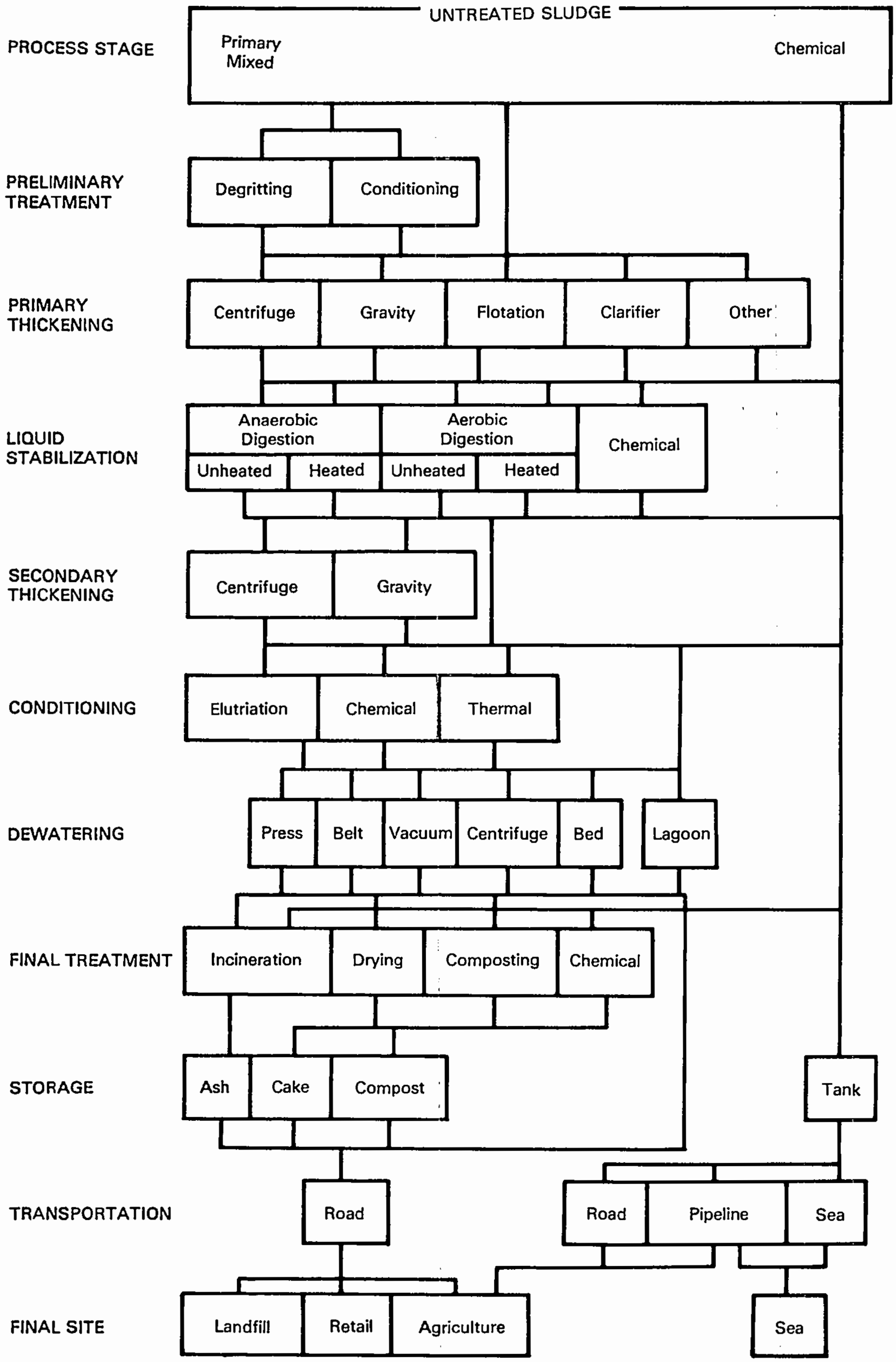
A dewatering process cannot be evaluated without considering the other processes involved in the overall wastewater/solids handling system. This evaluation or selection can be a complex procedure because of the large number of possible combinations of unit processes available for wastewater treatment and sludge thickening, stabilization, conditioning, dewatering, and ultimate use/disposal. Figure 2-1 shows the unit processes most commonly used to perform most of these

Table 2-2. Distribution of Sludge Disposal/Utilization Practices for 1,011 Surveyed POTWs (by % of dry sludge solids).

Practice	Small POTWs (<1 mgd)	Medium POTWs (1 - 10 mgd)	Large POTWs (>10 mgd)	Total of All POTWs
Landfill	31	34	12	15
Incineration	1	1	32	27
Land Application	39	38	21	24
Distribution and Marketing	11	17	19	18
Ocean Disposal	1	-	4	4
Other	17	10	12	12
	100	100	100	100

mgd x 0.0438 = m^3/s.

Figure 2-1. Sludge processing options.



functions. An evaluation procedure should start at the bottom of the figure with the use/disposal options and work back to a decision on the dewatering technology.

This chapter discusses regulatory concerns and the capabilities of mechanical dewatering processes.

2.2 Regulatory Concerns

Selection of dewatering equipment is seldom directly governed by regulations and/or guidelines. Reviewing authorities are concerned, however, with the equipment's performance capabilities, its reliability, and downtime for maintenance. Indirectly, the selection of dewatering equipment is influenced by the choice of a use/disposal system for sludge and the wastewater treatment system.

Before selecting a dewatering process, the design engineer should consult all regulations for a particular use/disposal system to see if a minimum cake solids concentration is required. For example, state landfilling regulations usually stipulate a required minimum cake solids concentration. This minimum level can vary widely from state to state.

2.3 General Performance Capabilities of Mechanical Dewatering Processes

In recent years there have been great advances in mechanical dewatering processes. To help the designer sort these processes out, this section of the manual compares the performance of several mechanical dewatering processes. However, communities with adequate land available should also study the air drying alternatives presented in Chapter 6, Air Drying Processes.

All of the various methods of mechanical sludge dewatering have the capability to produce good recovery (>90 percent) of feed solids and thus the major differentiation is the cake solids content. The capital cost and the O&M costs associated with dewatering may be of secondary concern if there is a high cost associated with the water content of the sludge.

The range of cake solids produced by a common type of dewatering process is the result of several factors. These factors are discussed in more detail in the specific sections dealing with each dewatering unit. However, the key factors are as follows:

a. The ratio of primary to secondary sludge. Inherently, secondary sludge will retain at least twice as much water, kg H_2O/kg total solids (TS), as primary sludge.

b. The origin of the secondary sludge. High SRT (sludge residence time) sludge retains more water than low SRT sludge. Bulking sludges will retain more water than non-bulking sludges.

c. The type and quantity of chemical conditioning can either enhance or reduce the cake solids, depending on the dewatering process employed.

d. The design and age of the dewatering equipment. Older equipment cannot compete with modern (1980 and newer) designs. Further, there may be several models of the same type of equipment, some of which will produce a drier cake than others. Good examples are the various belt press offerings and the diaphragm recessed plate press vs. the standard recessed plate press.

e. The design and operation of the dewatering stations will significantly impact cake solids content. Dewatering equipment operated at maximum solids capacity may sacrifice 3-5 percentage points in the final product dryness. Drier cakes are produced at reduced operating loadings.

f. Industrial discharges can both enhance or detract from the capability of a dewatering unit. Fibrous discharges, for example, can result in belt presses producing a much drier cake.

Figures 2-2, 2-3, and 2-4 represent typical ranges of cake solids produced by various means of mechanical dewatering. This data is representative of both organic (polymer) and inorganic conditioning; cake solids are not adjusted for ferric chloride and lime addition in these figures.

The figures are not recommended or suggested for design purposes and are provided only to indicate the relative capabilities of each type of dewatering unit and the effect of various types of sludges on cake solids. Depending on specific site factors, better or worse results can be achieved.

In Figures 2-2, 2-3, and 2-4, ferric chloride and lime are generally employed for vacuum filters and filter presses. While analyses of cake solids indicate that these cakes are higher in percent TS by 2-5 percentage points than a polymer conditioned sludge from the same device, the water content per unit weight of sludge solids is about the same. That is, a ferric chloride and lime vacuum filter cake of 23 percent TS would contain the same weight of water/kg sludge as would a polymer conditioned cake of 20 percent TS.

Figure 2-2 shows the characteristic results achieved when dewatering 100 percent raw primary sludge (RPS) and 100 percent waste activated sludge (WAS) using various kinds of dewatering equipment. If the sludges were dewatered first and mixed after

Figure 2-2. Dewatered sludge cake percent solids for raw primary and raw waste activated sludges.

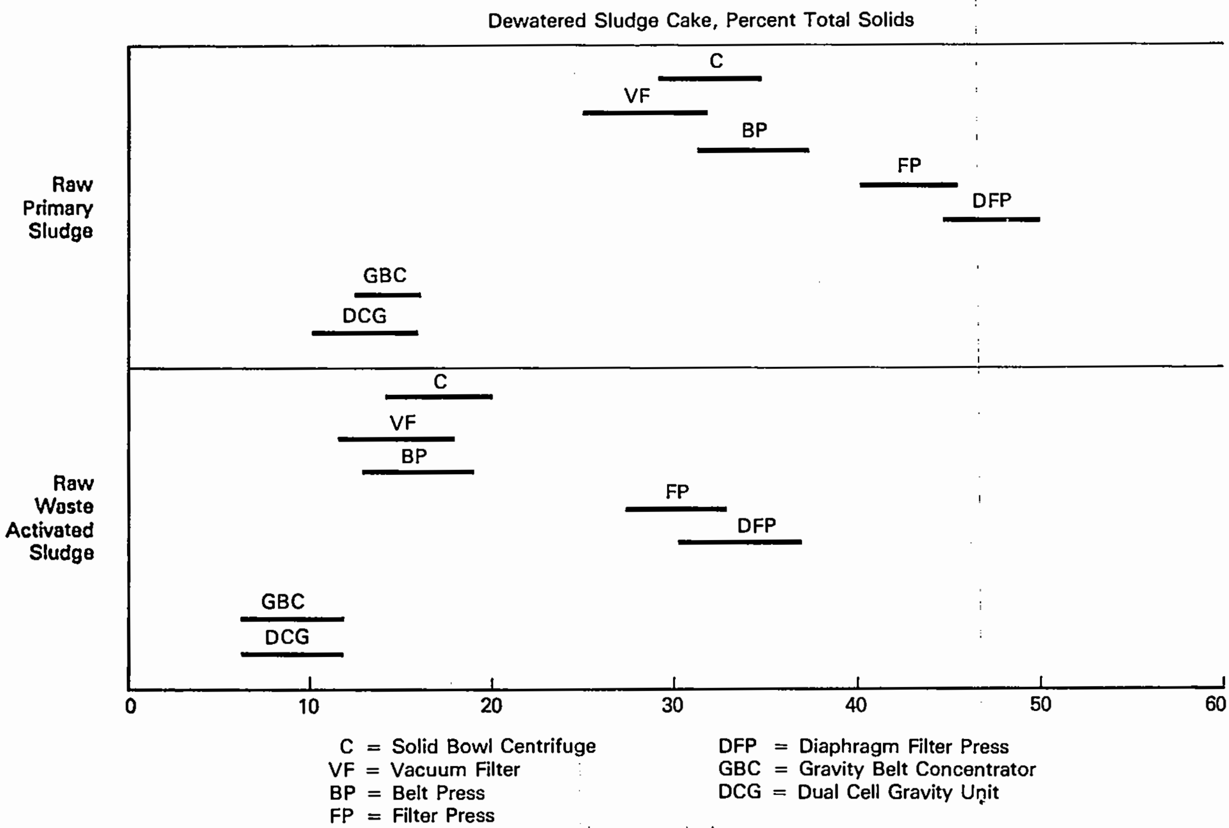


Figure 2-3. Dewatered sludge cake percent solids for mixtures of digested primary (P) and digested waste activated sludges (WAS).

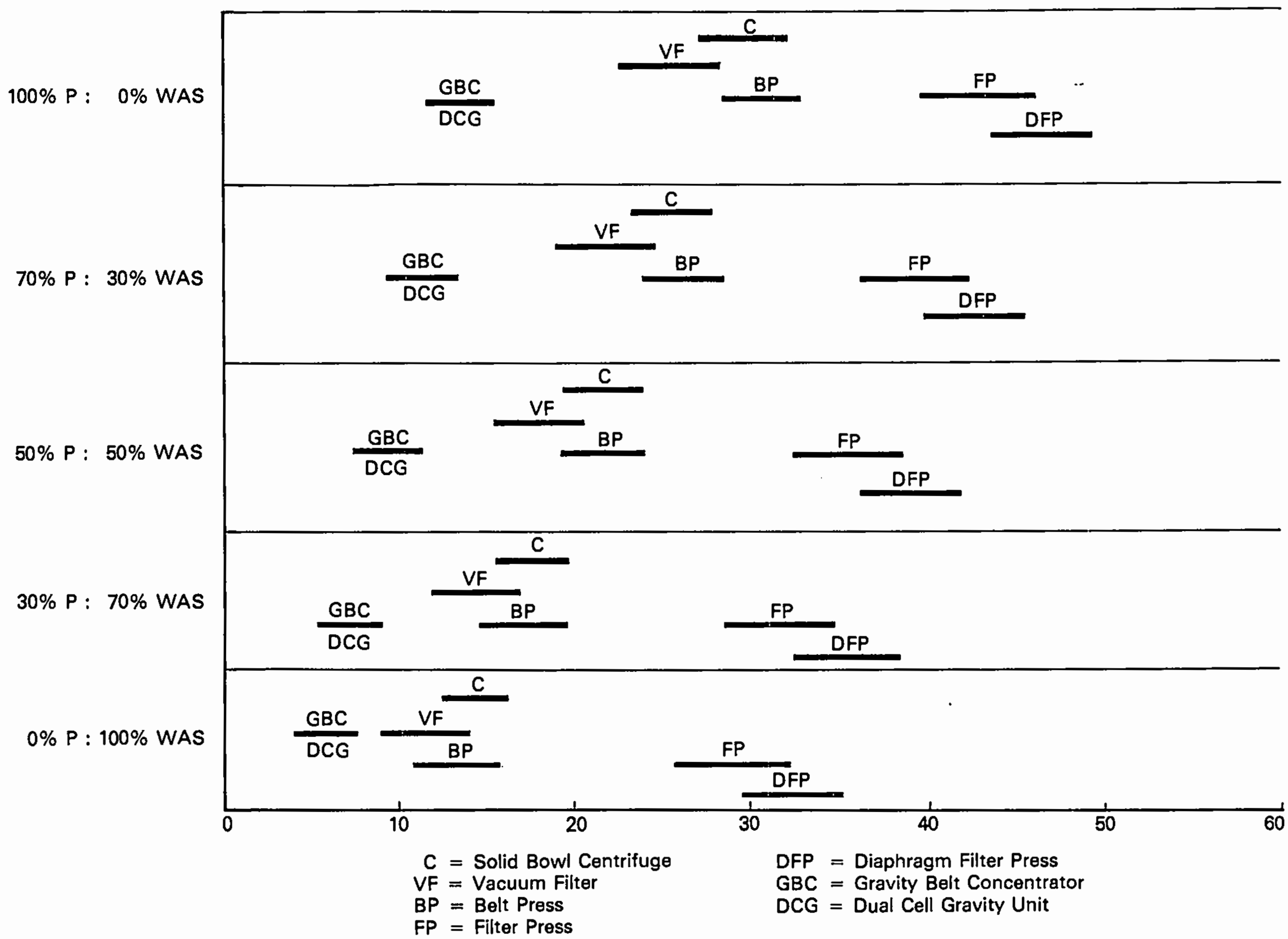
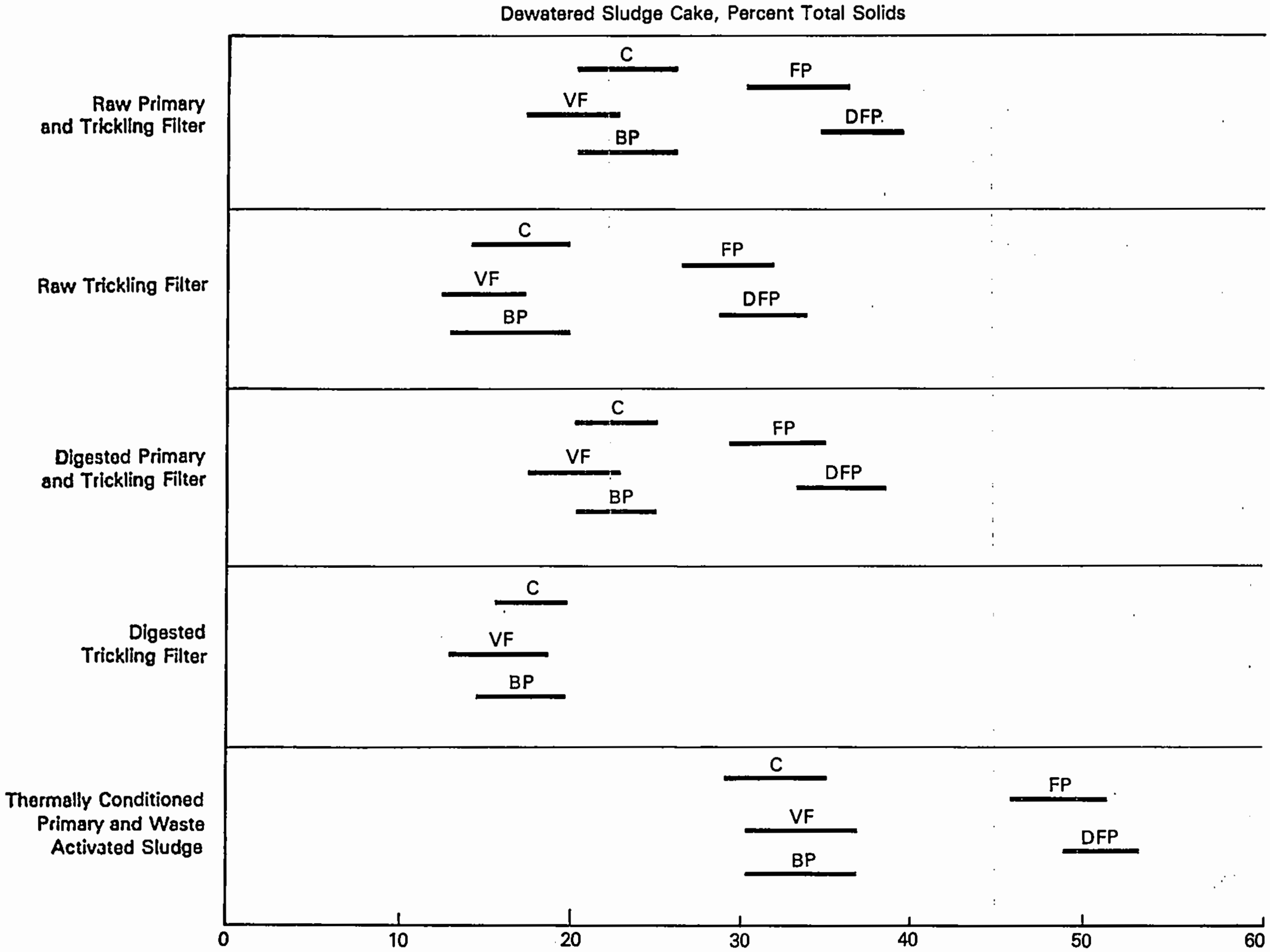


Figure 2-4. Dewatered sludge cake percent solids for mixtures of raw and digested primary and secondary sludges and heat-treated primary and secondary sludges.



C = Solid Bowl Centrifuge
VF = Vacuum Filter
BP = Belt Press
FP = Filter Press
DFP = Diaphragm Filter Press

dewatering, the final moisture content of the cake could be calculated from this figure as shown below:

Sludge at 60:40 RPS:RWAS

$$\%\,TS\,Mixture = \frac{(P + WAS)}{\dfrac{P}{\%TS_P} + \dfrac{P}{\%TS_{WAS}}} \qquad (2-1)$$

$$\%\,TS\,Mixture = \frac{(60 + 40)}{\dfrac{60}{30} + \dfrac{40}{17}}$$

$$\%\,TS\,Mixture = 23\%\,TS$$

It is necessary to substitute site-specific numbers for the sludge proportion of 60:40 and cake solids for each fraction. These numbers must be consistent with the sludge at the specific site and representative of the capabilities of the equipment supply and process design. This equation can also be used to estimate the moisture content of dewatered cake if the mixing is done before dewatering. The estimate is not conservative - it would be best to carry out dewatering tests on the mixture.

The filter presses (recessed plate and diaphragm plate) will always produce the driest cake solids with the available technology. New dewatering equipment, now in development or an early stage of application, may rival the filter presses' ability to produce the driest cake solids. Belt presses and centrifuges both achieve approximately the same cake solids, with the belt press more appropriate for sludges of higher primary (more structured) content when highest solids content is required.

Figure 2-3 provides cake solids content for various methods of dewatering a digested primary and waste activated sludge. Generally, digesting the primary sludge before dewatering results in a slightly wetter cake, probably a result of the finer particles produced by anaerobic decomposition. However, those plants with poor grit removal, and hence a higher proportion of grit in the digested sludge, may produce a drier digested sludge than would be produced if it were dewatered in the raw state.

Mechanical thickening devices such as the Gravity Belt Concentrator and the Dual Cell Gravity Unit produce a much lower solids content then the other dewatering devices. However, their simplicity and ease of operation can be well suited to small plants that either stockpile and/or land spread dewatered sludge.

Figure 2-4 provides dewatering information for raw and digested trickling filter plant sludges. These sludges behave similarly to an activated sludge plant's waste solids where the sludge ratio is about 65:35 RPS:WAS.

Heat-treated sludges are often dewatered without chemical conditioners. However, they will require polymers to control the suspended solids recycle with centrifuges and belt presses. However, dosage will be low - about 20-25 percent of that required for the sludge before thermal treatment.

Some general conclusions can be drawn from these figures. They are:

a. Solid bowl centrifuges and belt presses can produce about the same cake solids. However, belt presses with high pressure attachments can generally produce 2-3 percentage points higher cake solids with sludges of good structural characteristics.

b. The low energy gravity drainage units that produce a cake of a low solids content may be attractive for small plants and where land spreading disposal is practiced.

c. The recessed plate filter presses will produce 6-10 percentage points drier cake than the continuously fed dewatering units; addition of the diaphragm can increase the cake solids content 3-5 percentage points more.

d. In general, digested sludge cakes will have a solids content 2-3 percentage points lower than that of the raw sludge, with the exception of those with a low volatile content due to grit.

e. The range in cake solids from plant to plant will be quite wide due to specific site conditions and design and operating factors.

2.4 Key Operations Variables Affecting Mechanical Dewatering Performance

Table 2-3 presents the more significant design and operating variables that impact the performance of the specific dewatering device. (More detailed information is provided in the various sections of the manual reviewing each dewatering process.) These variables affect the rate of production, solids recovery, and cake solids, as well as the type and quantity of conditioning chemicals and dewatering aids employed.

Table 2-3 lists only those variables that affect performance and that can be changed in the field and during design of the dewatering station. Variables

Table 2-3. **Operational Variables for Mechanical Dewatering Processes**

Vacuum Filter	Recessed Plate
1 Drum Speed/Cycle Time	1 Pressure of Feed Sludge
1 Drum Submergence	1 Filtration Time
1 Sludge Feed Concentration	1 Conditioning Chemicals
2 Quantity of Wash Water Used	2 Use of Precoat
2 Filter Media Used	2 Frequency of Cloth Washing
2 Conditioning Chemicals - Type and Dosage	2 Filter Cloth Used
3 Vat Agitation	
Belt Filter Press	Diaphragm Plate Press
1 Belt Speed	1 Diaphragm Pressure
1 Sludge Feed Concentration	1 Conditioning Chemicals
1 Polymer Conditioner	• Type & Dosage
• Type & Dosage	• Point of Addition,
• Dosage	2 Pressure of Feed Sludge
• Point of Addition, Contact Time, Mixing	2 Filtration Time
	2 Diaphragm Squeezing Time
2 Belt Tension	2 Filter Cloth Used
2 Belt Type	2 Frequency of Cloth Washing
2 Washwater Flow & Pressure	2 Sludge Feed Concentration
Solid Bowl Centrifuge	Gravity/Low Pressure Dewatering
1 Bowl/Scroll Differential Speed	1 Polymer Dosage
1 Pool Depth	1 Retention Time
2 Polymer Conditioner	2 Sludge Feed Concentration
• Dosage	2 Belt Speed
• Point of Addition	2 Force Applied By Rollers
2 Mode of Differential Speed Control	2 Depth of Dewatered Sludge in Cylindrical Devices
3 Sludge Feed Concentration	

Legend: 1 Very important variable
2 Significant variable
3 Less important variable

controlled by the equipment suppliers, such as the bowl angle of a centrifuge, have not been included.

The table does not include sludge feed rate. If you set the variables shown for a vacuum filter, feed rate is fixed. The same is true for pressure filters. In a belt press, increasing feed rate will flood or starve the filter. In a centrifuge, you can vary feed rate at will - you will just get a more or less dry cake and less or more solids in the centrate. Feed solids concentration has been included since some dewatering processes are more sensitive to feed concentration than others.

Chapter 3
Sludge Characteristics and Preparatory Treatment

3.1 Sludge Production and Concentration

3.1.1 Introduction

There are several sources of wastewater sludges; these sludges can vary widely in characteristics and quantity. From the standpoint of quantity per unit of flow, the principle variables are the strength of the wastewater, whether chemicals are utilized in the process, and the degree of treatment.

The typical wastewater sludges are classified as primary, biological, and chemical. The biological sludges produced are activated sludge and fixed film sludges from rotating biological contactors and trickling filters. The activated sludge may have primary sludge solids incorporated into the biomass when primary clarifiers are not employed.

Chemical sludges may be produced simultaneously with primary sludge or biological sludge through the addition of metal salts for precipitation of phosphorus, or they can be made in a separate tertiary treatment stage. Lime is sometimes used in the primary treatment stage and also in a tertiary stage, when softening of the effluent is required for reuse. The reader is also referred to the EPA Technology Transfer *Process Design Manual - Phosphorus Removal* (EPA-625/1-87/001) for a discussion of the production and dewatering characteristics of chemical sludge.

In some cases, well designed sludge handling systems were actually marginal in operation due to inaccurate estimates of wastewater treatment loadings of BOD_5 (5-day biochemical oxygen demand) and TSS (total suspended solids) and the subsequent sludge production. These problems occurred for a variety of reasons as outlined below:

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

3.1.2 Primary Sludge

The raw primary sludge (RPS) production is easily determined from the total flow and the influent and effluent TSS (total suspended solids) of the primary clarifier. Care should be taken to ensure that the influent sample, which does not contain recycled solids, is the same as the primary clarifier influent. Some adjustment of the influent is necessary to account for recycled solids removed by primary clarification. Even with good operation, recycled TSS can amount to 15-20 percent of influent TSS, and the BOD_5 recycle is usually 8-15 percent of the influent BOD_5.

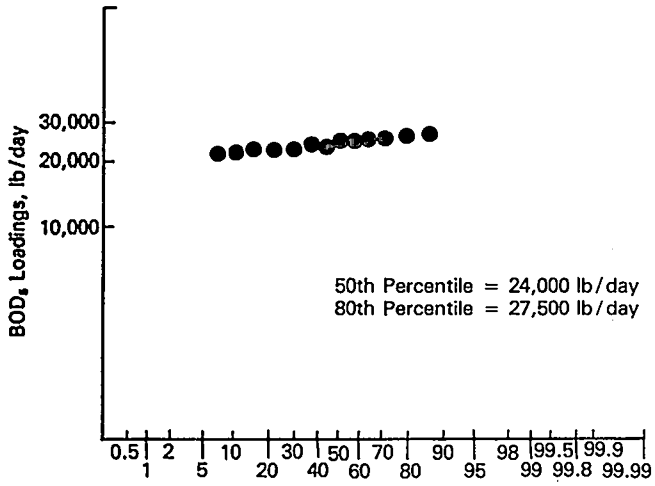
The influent loadings and resulting sludge production should be analyzed and developed into a frequency plot, which would indicate the frequency of a specific TSS and BOD_5 influent loading (kg/d) vs. $\leq$ % time (frequency). Similar graphs should be plotted for RPS produced and PE (primary effluent) BOD_5 (kg/d vs. % time). Figure 3-1 is a typical example.

A mass balance of the overall process should be prepared to ensure accounting of all TSS and BOD_5. An example is shown in Figure 3-2, and indicates the importance of identifying the magnitude of the recycle BOD_5 and TSS.

Unlike secondary sludge, the volatility of primary sludge may vary considerably from day to day and seasonally. This is particularly true of sewage systems with combined sewers and/or substantial infiltration and inflow.

The domestic and commercial discharges of volatile suspended solids (VSS) would not vary widely throughout the year. Some short-term increases may be noted due to "first-flush" effects during sudden wet weather conditions. First-flush effects

Figure 3-1. Influent BOD5 average monthly loading, 1985-1986.



lb/day x 0.4536 = kg/day.

occur with the transport of accumulated solids in the sewers and street washing where there are combined sewers. Where there are seasonal or variable industrial discharges, the VSS may vary widely, and the primary removals of TSS and BOD_5 may also vary, depending on the nature of the solids.

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

	TSS, lb/d	VSS, lb/d	Volatile %
Dry Weather	100,000	78,000	78
Wet Weather	130,000	78,000	60

lb/d x 0.454 = kg/d

The sludge production at the lower volatile content is 30 percent higher than the dry weather sludge quantity. While the lower volatile content sludge is somewhat easier to handle due to the grit content, the sludge handling design should anticipate the higher quantity of sludge solids. In existing plants, the past operating records should be scrutinized to determine if there is a significant variation in the actual sludge volatility and the sludge quantity projected.

The efficiency of primary clarification is important because not only is the primary sludge easier to handle, but the unit yield (kg/kg) of secondary sludge is partially dependent on the TSS/BOD_5 in the clarified mixture. Figure 3-3 presents the yield and the dewatered cake concentration of various RPS:WAS ratios. When primary clarifiers are not employed, the total quantity of solids produced is lower than RPS + WAS, but the water retention characteristics of the biological solids increase. While the absolute values shown in Figure 3-3 vary, depending on sludge characteristics and the mechanical equipment employed, the general relationship holds.

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

ENGLISH	OFR, max $\leq$ 10 SWD^2	(SWD = 6 $\leq$ 10 ft)
	OFR, max $\leq$ 100 SWD	(SWD = 10 $\leq$ 15 ft)
METRIC	OFR, max $\leq$ 4.5 SWD^2	(SWD = 2 $\leq$ 3 m)
	OFR, max $\leq$ 12.25 SWD	(SWD = 3 $\leq$ 5 m)

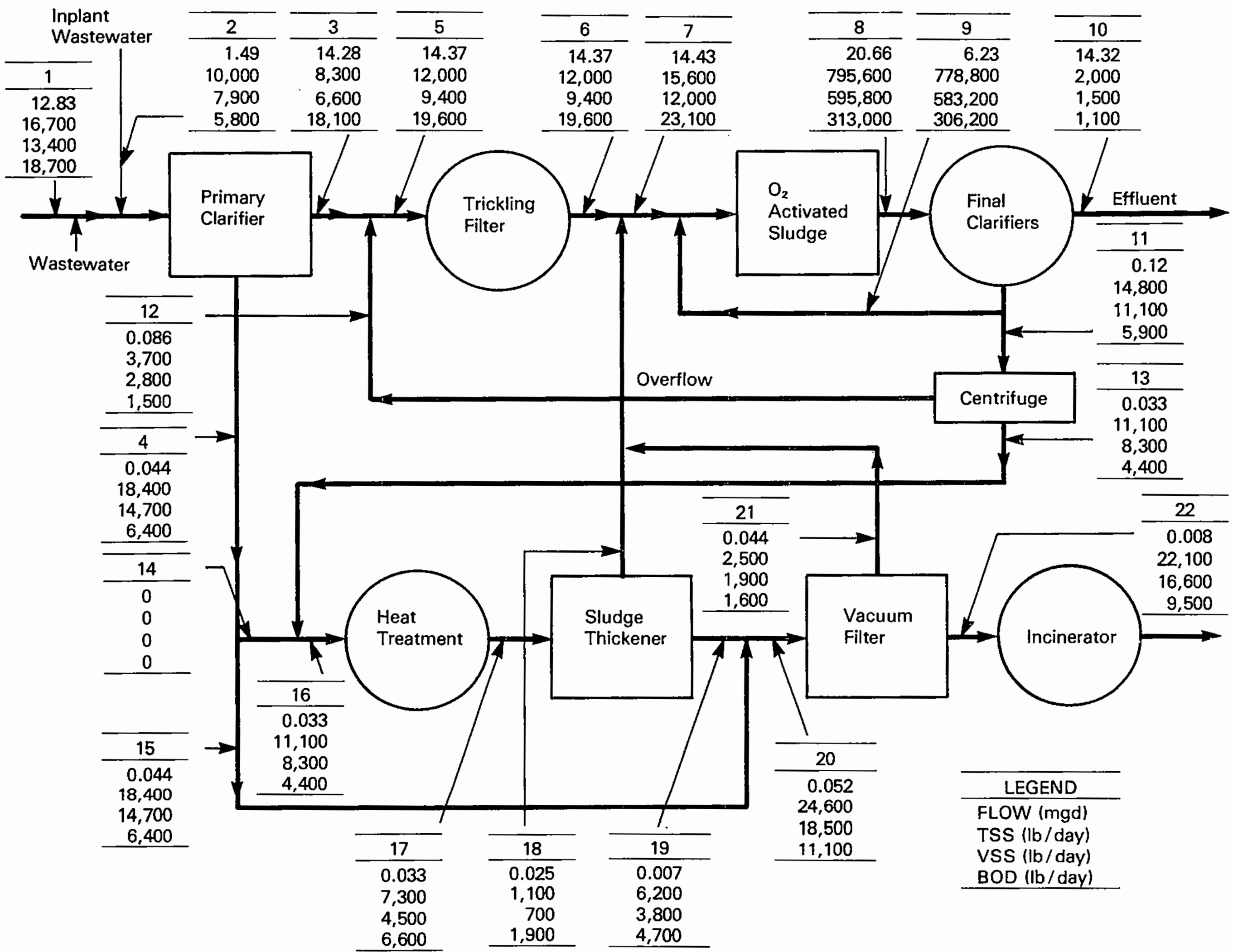
For rectangular clarifiers, the length of the flow path is most important to overcome inlet disturbances. The depth of the basin is also significant. For basins less than 30 m long, the length to width ratio is $\geq$5:1. Basins that are 30-65 m long and 4-5 m deep will provide excellent results, even at rates up to and exceeding 67.2 $m^3/m^2/d$ (1,650 gal/sq ft/d).

3.1.3 Chemical Treatment of Raw Wastewater

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

As discussed in Section 5.5.3.1, Table 3-1 provides estimated quantities of suspended and chemical solids removed in a hypothetical primary sedimentation tank processing wastewater that has been treated with lime, aluminum sulfate, or ferric chloride. The use of polyelectrolytes may greatly enhance the solids capture in the clarifier. The removal of TSS is usually in the range of 75-85 percent and BOD_5 removal is 55-70 percent depending on the specific wastewater characteristics.

Figure 3-2. Process mass balance.



The chemically defined soluble BOD_5 will be appreciably less than the filtrate soluble BOD_5 since some colloidal BOD_5 will be agglomerated and settled.

The solids precipitated are defined as $Ca_x(PO4)_y$, $CaCO_3$, $FePO_4$, $Fe(OH)_3$, $AlPO_4$ and $Al(OH_2)_3$. However, some of these precipitates will be hydrated and represent more sludge than shown in Table 3-1. This is why the metal salt precipitates are so voluminous. When sludge solids are dried at 103-105°C (217-221°F) for solids analysis, some of this water is lost during the test procedure. However, the hydrate moisture adversely affects the ability of the sludge to be thickened and dewatered.

3.1.4 Biological Sludge Yield

Sludge yields will vary widely from plant to plant depending on the overall treatment plant configuration, wastewater characteristics, and the biological kinetics/parameters employed for design. A wastewater with a high COD/BOD_5 ratio will produce more excess biological sludge solids. Net yields (Y_N) are determined from Equations 3-1 and 3-2.

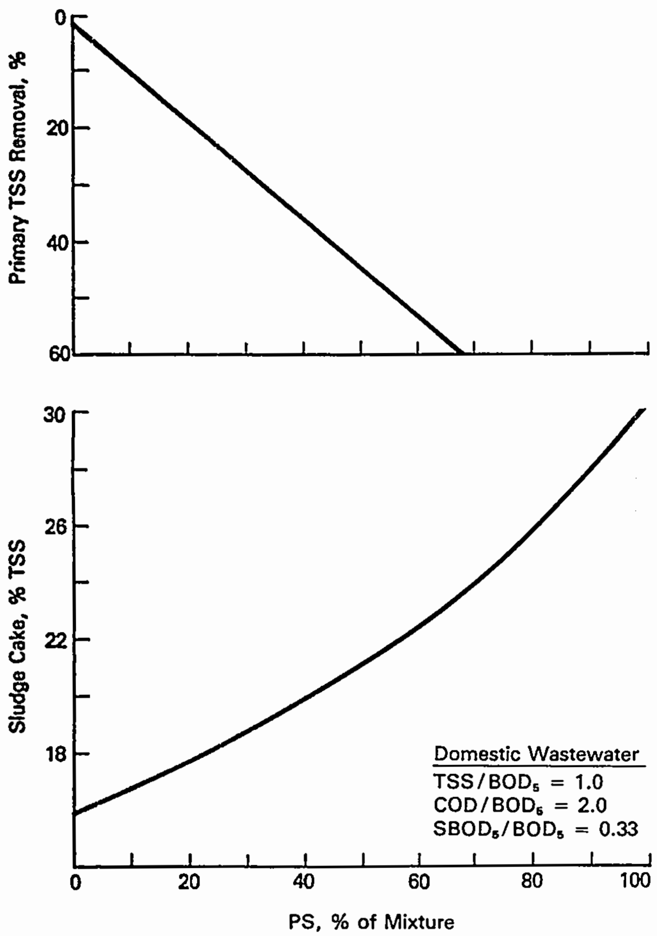
$$Y_N = a\ BOD_{5R} - b\ (M) \qquad (3\text{-}1)$$

$$1/SRT = a\ (F_R/M) - b \qquad (3\text{-}2)$$

where,

SRT = M/Y_N
SRT = aerobic sludge age, days
F_R = BOD_{5R} (removed), kg/d
M = MLSS or MLVSS, kg
a = synthesis, kg/kg BOD_{5R}
b = endogenous decay, day^{-1}
Y_N = net yield, kg TSS or kg VSS

Figure 3-3. Cake solids as a function of primary clarifier efficiency and PS:WAS ratio.



Both the synthesis value "a" and endogenous decay value "b" are reported in the literature on a TSS and VSS basis and thus the origin of these values must be distinguished. As shown in the WPCF Manual of Practice No. 8, *Treatment Plant Design*, the COD/BOD_5 and TSS/BOD_5 values must be specified in order to project a net yield as a function of SRT and temperature. Thus, the values of "a" and "b" are variable and both temperature and SRT dependent. Since sludge yields are higher at lower temperatures, it is necessary to use the colder period of the year to project maximum sludge production, when the organic loading is uniform throughout the year.

Figure 3-4 indicates the range of sludge production to be expected at temperatures between 10 and 30°C (50-86°F) as a function of the SRT with and without primary treatment. In this case, the wastewater is specifically identified as typical domestic wastewater at 400 mg/l COD, 200 mg/l BOD_5, and TSS, and primary clarifier effluent is as noted. This curve should not be used for wastewaters having different relative $COD/BOD_5/TSS$ ratios either in the raw wastewater or in the settled primary effluent.

When the COD/BOD_5 ratio of the primary effluent exceeds the values shown in Figure 3-4, higher excess biological solids production will often occur if the COD is removable by adsorption/oxidation. Los Angeles-Hyperion, Columbus-Southerly, and Columbus-Jackson Pike all have primary effluents of 2.3-2.7:1 COD/BOD_5. Secondary sludge yields of 0.75-0.90 kg EAS/kg BOD_5 are produced (EAS = excess activated sludge (WAS + effluent TSS)]. Not suprisingly, the COD yields (kg EAS/kg COD_R) are more consistent. The BOD test may be, on occasion, a poor indicator of the yield.

3.1.5 Biological Phosphorus Yield

When phosphorus is removed, the net sludge yield will increase measurably. The increase will be similar to that experienced when Fe^{+++} or Al^{+++} addition is employed to precipitate the phosphorus chemically. That is, the excess phosphorus precipitated by manipulation of the biological environment is also an inorganic salt of K, Ca and Mg. It is recommended that the following procedure be used to determine this excess sludge.

Influent BOD_5:	140 mg/l
Effluent $SBOD_5$:	5 mg/l
$\therefore BOD_{5R}$:	135 mg/l
Influent TP:	8 mg/l to bio-treatment
Effluent STP:	2 mg/l (soluble)
Calculated sludge yield:	65 mg/l VSS 80 mg/l TSS
Normal sludge P:	2.0 percent VSS
Excess bio-P removal	= 8 - 2 - 0.02(65) = 4.7 mg/l P

Additional sludge = 4.5 x 4.7 = 21 mg/l
(4.5 mg/mg ΔP)

Total sludge production = 80 + 21 = 101 mg/l

The value of 4.5 was based on an average MW of 140 for the inorganic phosphorus crystals in the biological cells. In this example, the biological sludge yield would have increased from a calculated value of 0.61 mg EAS/mg BOD_{5R} to 0.75 mg/mg, or 21 percent higher.

Typical yield coefficients found in all textbooks and other reference materials do not allow for this higher sludge yield. However, recent papers in biological phosphorus removal have confirmed the MW is about 140.

The phosphorus balance [primary effluent (PE) to final effluent (FE)] provides an easy and direct method to determine the net yield, SRT, and system MCRT (mean cell residence time). MCRT in this

Table 3-1. Typical Production of Primary and Primary-Chemical Sludges[1]

Mode of Operation	Dosage of Chemical	Raw TSS Removed	Raw BOD_5 Removed	Chemical Sludge Produced
	mg/l	mg/l	mg/l	mg/l
Plain Sedimentation	-	120	60	-
Polymer Added	0.5 - 3.0	150	90	-
CaO Aided[2]	200	160	120	128
$FeCl_3$ (as Fe)[3]	12	160	120	47
Al_2SO_4 (as Al)[3]	12	160	120	46

[1] Based on 200 mg/l BOD_5, 200 mg/l TSS, and 10 mg/l TP in raw sewage; primary effluent ≤ 2 mg/l total phosphorus.
[2] Varies due to permanent hardness in the water, used 35 mg/l precipitated as $CaCO_3$.
[3] May require polymer addition to enhance clarification.

discussion includes the clarifier sludge inventory. Equation 3-3 describes the procedure.

$$SRT = \frac{(MLSS - kg)(\% P/100)}{(PETP - FETP)(m^3/d)(1{,}000)} \qquad (3-3)$$

$$MCRT = \frac{MLSS + Clarifier\,TSS}{MLSS}(SRT)$$

$$Sludge\,Yield = \frac{MLSS}{SRT} \text{ or } \frac{MLSS + Clarifier\,TSS}{MCRT}$$

The method is quite accurate if phosphorus removal is relatively constant and the ratio of MLSS to the clarifier sludge is relatively constant. Three- or five-day running averages are better estimates of Y_N, SRT, and MCRT.

3.1.6 Chemical Phosphorus Precipitation in the Biological System

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

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

$AlPO_4$ = 121/31 or 3.9 mg/mg P removed
$Al(OH)_3$ = 77/26 or 3.0 mg/mg excess Al
$FePO_4$ = 151/31 or 4.9 mg/mg P removed
$Fe(OH)_3$ = 107/56 or 1.9 mg/mg excess Fe

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

Molar Ratio (metal:TP)	Residual STP (mg/l)
1.0	2
1.5	1
2.0	0.3

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

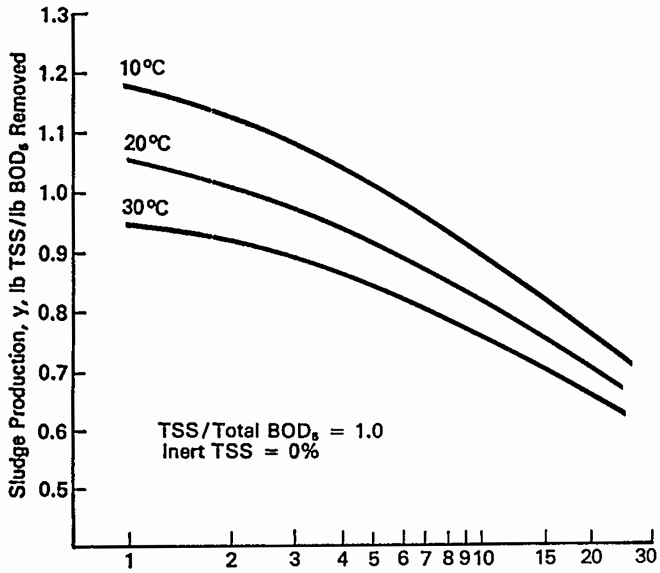
3.2 Sludge Concentration - Primary Clarifiers

Sludge feed concentration is an important factor in sludge processing units such as digestion and dewatering. In larger plants separate sludge pre-thickening is economically viable, but this is not true for most plants of less than 0.2 m^3/s (5 mgd). [Over 95 percent of the sewage treatment plants in the United States are less than 0.2 m^3/s (5 mgd)]. While it is not a specific goal of this manual to provide information regarding pre-concentration of sludges, it is necessary to consider the impact of sludge concentration on sludge dewatering and sludge handling in general.

Most of the smaller plants in the United States will not have separate thickening and will utilize primary treatment units to co-settle and thicken raw primary and waste secondary solids. The problems encountered in this procedure are well known and documented. Many of the problems associated with using the primary clarifier as a thickener are the result of excessive solids retention time in the clarifier. It is sludge retention time, not liquid retention time, that is the primary cause of odors in primary clarifiers. Such odors result from increasing the sludge blanket (inventory) over the sludge withdrawal pipe to maximize the sludge concentration and prevent rat-holing of the sludge. It is, however, possible to use

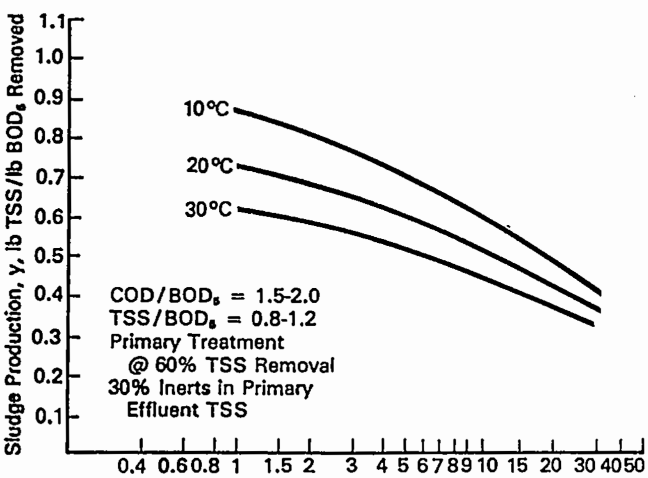
Figure 3-4. Process mass balance.

Sludge Yield Without Primary Clarification



Solids Retention Time, days

Sludge Yield With Primary Clarification



Solids Retention Time, days

this procedure effectively with minor changes in the design approach and negligible additional costs.

Where the primary clarifier must serve a dual function of clarifying the wastewater as well as delivering a concentrated sludge for digestion or dewatering, the conventional primary clarifier design configuration is inappropriate. This is particularly true for primary clarifiers in smaller plants, where it may be necessary to have 1.5-2.0 days SRT in the clarifier to create a 1.0 to 1.5 m (3-5 ft) sludge blanket above the sludge withdrawal pipe. The build-up of sludge to produce a thicker underflow interferes with clarification (lower efficiency) and sometimes results in gasification, odors, and floating sludges.

However, in the new plants there is an easy remedy for this problem. Construct the smaller clarifiers with the standard thickener floor slope of 2.75:12. In larger clarifiers, use a dual slope clarifier where the inner slope of 2.75:12 is the thickening zone and the outer zone is 1:12. On a primary clarifier, only 40-50 percent of the diameter would be required for sludge thickening; thus it is sufficient to modify the floor slope at mid-radius.

Figure 3-5 shows the types of clarifier floor configurations of which only three are suitable for efficient combined clarification and thickening. Type A is the design most commonly employed, but it is not suitable for combined clarification and thickening. Types B, C and D all can provide much better performance in terms of thickened sludge concentrations and lowest sludge inventory; hence, they provide freshest sludge and highest flexibility in terms of sludge removal and ease of operation. Pertinent dimensions for a 30 ft diameter and 80 ft diameter primary clarifier-thickener are also shown in Table 3-2.

Process data used to construct the sludge level in Table 3-2 are provided below. Underflow concentrations were based on general experiences with sludge thickening.

RPS @ 120 mg/l	=	1,000 lb TSS/mgd
WAS @ 80 mg/l	=	667 lb TSS/mgd
Total	=	1,667 lb TSS/mgd

9.1 m (30 ft) diameter @ 1.36 m^3/m^2d (800 gal/sq ft/d):

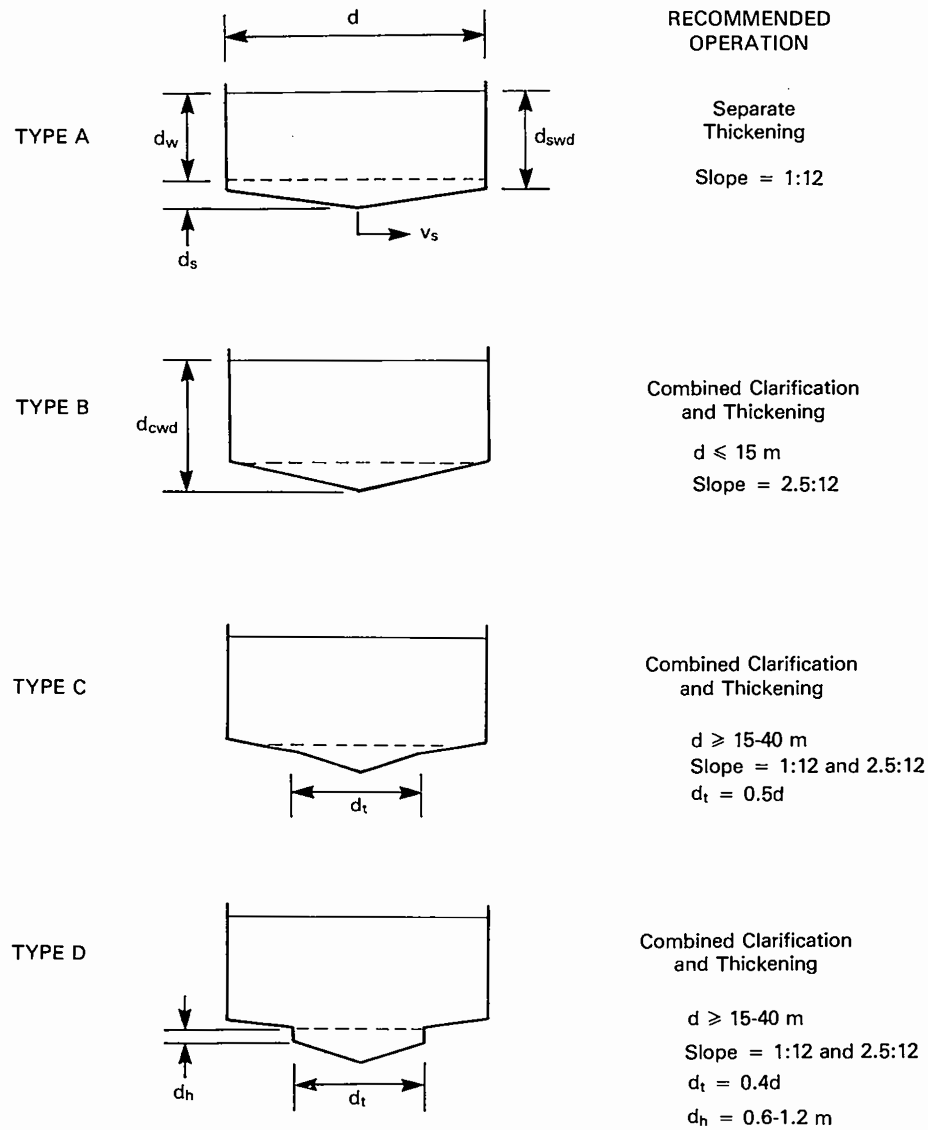
TSS_R (TSS removed) = 943 lb TSS/d

24 m (80 ft) diameter @ 32.6 m^3/m^2d (800 gal/sq ft/d):

TSS_R = 6,706 lb TSS/d

Sludge blanket depth in small clarifiers is a two-fold problem in units with the 1:12 floor slope. As shown in Table 3-2, the 1.0 day SRT sludge depth is only

Figure 3-5. Effect of floor configuration on blanket location and sludge inventory.



LEGEND

d = Diameter
d_w = Depth of Water
d_s = Depth of Sludge
d_{swd} = Side Water Depth
d_{cwd} = Center Water Depth
d_t = Diameter at Steeper Slope
d_h = Height of Vertical Wall
v_s

Table 3-2. **Primary Clarifier Comparison - Sludge Blanket Location, and Sludge Concentration vs. Floor Configuration (see Figure 3-5 for nomenclature)**

	Type			
	A	B	C	D
Slope	1:12	≥ 2.5:12	1:12 ≥ 2.5:12	1:12 ≥ 2.5:12
30-ft diameter				
d_{swd}, ft	9.0	9.0	NR	9.0
s_{cwd}, ft	10.3	12.1		14.1
d_s[1], ft	1.9	2.9		4.3
d_w, ft	8.4	9.0+		9.0+
d_h, ft	-	-		3.0
V_C, ft^3	313	737		612[2]
V_S, ft^3/d	544	428		313
Underflow, % TSS	2.75	3.5		4.5
80-ft diameter				
d_{swd}, ft	11.0	NR	11.0	11.0
s_{cwd}, ft	14.3		16.8	18.8
d_s[1], ft	3.2		5.0	6.2
d_w, ft	11.0+		11.0+	11.0+
d_h, ft	-		-	2.6
V_C, ft^3	5,582		6,623	7,726
V_S, ft^3/d	3,546		2,680	2,364
Underflow, % TSS	3.0		4.0	4.5

[1] Based on 1,000 lb PS/d + 600 lb WAS/mgd sludge inventory @ 0.75 % TSS in underflow (1-day SRT).
[2] Based on conical plus cylindrical volume below cone.
NR = not recommended.
V_C = volume of conical section
V_S = volume of sludge at 1-day inventory.
ft x 0.03048 = m.
cu ft x 0.0283 = m^3.

0.58 m (1.9 ft) in clarifier type A. In smaller plants, normal practice would be to pump a fewer number of hours/day at a proportionally higher rate. This procedure generally results in a diluted sludge being processed with less efficiency.

Type D in Figure 3-5 will provide the highest sludge concentration since it most closely approximates a thickener design, has the deepest sludge inventory and has the sludge inventory closest to the withdrawal point. This design has been available and employed in both municipal and industrial facilities for wastewater treatment. The Type D unit is illustrated in Figure 3-6 and, as shown in Figure 3-5, the sludge blanket would still be in the lower zone with a sludge SRT of over one day in the unit. The slope is 0.5 to 1.0:12 in the outer area and 2.5 to 3.0:12 in the thickening zone.

Type B is a design that is simpler and quite suitable for clarifiers up to about 15 m (50 ft) diameter. The bottom slope would be 2.5 to 3.0:12 with the steeper slopes oriented to the smaller diameter. As a general rule of thumb, the depth (d_C) of the coned section should not be less than 1 m (3 ft). Beyond 15-18 m (50-60 ft) diameter, the depth of reexcavation becomes a factor and much of the coned volume is not needed for sludge storage.

Type C units are most often installed in large industrial clarification-thickening operations for both process and wastewater treatment. The configuration has been employed for both primary and secondary treatment stages. As in Type D, the slope of the outer zone is usually decreased as the diameter increases, since sludge conveyance to maintain inventory and depth are only critical in the thickening zone. Outer zone floor slopes of 0.5:12 are common above 30 m (100 ft) diameter to minimize the overall depth of the unit. While Type C and D clarifiers larger than 60 m (200 ft) in diameter have been installed, most municipal wastewater treatment plants with more than 15,000 sq ft of clarifier capacity will employ separate thickening.

Those smaller treatment facilities (less than 5 mgd) employing primary clarifiers may also find it advantageous to use a floor slope of 2.0 to 2.5:12 on secondary clarifiers. This slope will reduce the operational problems associated with maintaining a high MLSS (mixed liquor suspended solids) since there will be a higher return sludge concentration and lower waste sludge volume. The plant will also be more efficient and easier to operate, both in the wet and solids ends.

In larger plants without primary clarifiers, pre-concentration of the WAS is recommended prior to sludge dewatering for two reasons: (1) to reduce the volume; and (2) to provide some buffer capacity between the sludge wasting schedule and the dewatering operating schedule. The purpose of pre-concentration should be to increase the waste activated sludge (WAS) concentration by 50 to 100 percent, i.e., from 0.6-1.0 percent to 1.5-2.0 percent TSS. These high-rate gravity thickening systems can operate at 3 to 5 times the solids loading versus those where ultimate compaction is required. Flotation and centrifugal thickening are also employed to reduce waste sludge volumes prior to subsequent processing and dewatering and will produce 4-7% TS.

3.3 Characteristics of Waste Sludges

3.3.1 Specific Gravity and Volatility

The specific gravity of sludge will be in part a function of the amount of grit and fine inert particles in the sludge. These inorganic particles will have a specific gravity of 2.5-2.9. Where there is good degritting, the specific gravity of sludges will have the volatile and specific gravities shown in Table 3-3(2-4). The specific gravities of sludge solids are quite low and will vary depending on the source.

The specific gravity of fixed film biological sludge is generally higher than that of waste activated sludge.

Figure 3-6. Combination primary clarifier-thickener.

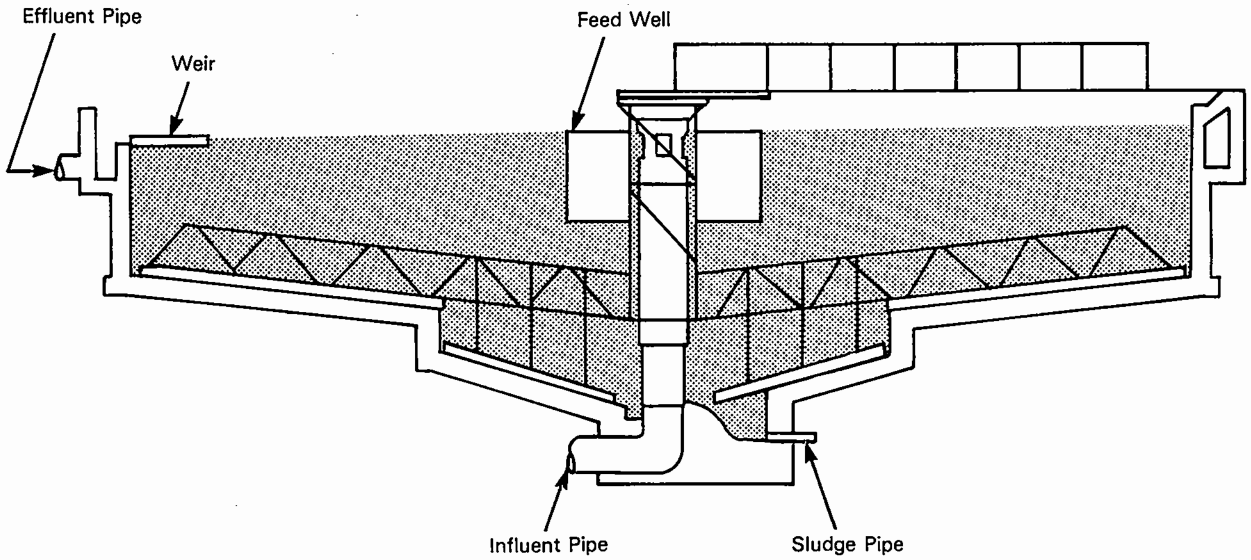


Table 3-3. Specific Gravity of Waste Sludges

Sludge Type	Volatility percent	Range of Specific Gravity g/cc
RPS	75 - 80	1 + 0.010 (%TSS) to 1 + 0.012 (%TSS)
WAS	80 - 85	1 + 0.007 (%TSS) to 1 + 0.012 (%TSS)
TF & RBC	75 - 80	1 + 0.015 (%TSS) to 1 + 0.025 (%TSS)
RPS + WAS	75 - 85	1 + 0.004 (%TSS) to 1 + 0.006 (%TSS)

RPS = Raw primary sludge.
WAS = Waste activated sludge.
TF = Trickling filter.
RBC = Rotating biological contactor.

This is evidenced by a lower SVI and generally higher settling rates. The specific gravity of the sludges after anaerobic digestion will increase due to reduction of some of the hydrous fractions and the increased inert content.

3.3.2 Pre-Concentration or Thickening of Waste Sludges

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

The general experience in thickening sludges is shown in Table 3-4. The results achievable in the primary clarifier are dependent on the clarifier design as reviewed earlier. Thickening increases the solids content of sludge slurry by a partial, but substantial, removal of the liquid phase. The purpose is to reduce the sludge volume to be stabilized, dewatered, or hauled away. Figure 3-7 shows the importance of thickening prior to mechanical dewatering. Thickening can be accomplished by partial thickening in a primary or secondary clarifier, a gravity thickener, a dissolved air flotation thickener, a centrifuge, a gravity or low pressure belt press, or a rotary drum device.

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

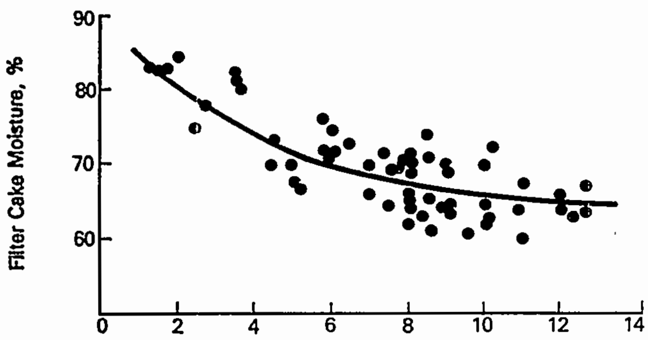
The use of gravity thickeners for both RPS + WAS has had mixed results. Most of the poor results can be traced to one or more of the following causes:

Table 3-4. Thickening of Waste Sludges

Type of Sludge	TSS Concentration, percent: Primary Clarifier	Flotation/DAF	Gravity Thickener	Belt Thickener[1]	Centrifuge
RPS	5 - 7	-	8 - 10	9 - 12	9 - 12
WAS		3 - 5	2 - 2.5	4 - 6	4 - 6
FFS[2]		3 - 5	2.5 - 3	5 - 7	5 - 7
RPS + WAS	2.5 - 4	4 - 6	4 - 5	5 - 7	5 - 7
RPS + FFS	3 - 5	4 - 6	5 - 6	5 - 10	6 -10

[1] Polymers required.
[2] Fixed film sludge.

Figure 3-7. Effect of feed solids on performance of a rotary vacuum filter.



a. RPS+WAS feed concentration is >0.5 percent TSS.

b. RPS is very septic.

c. WAS is > RPS fraction.

d. Secondary dilution water is inadequate.

e. Floor slope is <2.5:12, causing excessive solids retention.

f. Sludge is not removed continuously.

Properly designed and operated gravity thickeners work effectively on mixtures of RPS and WAS throughout the United States. Misusing them as sludge storage zones causes operator grief. If storage is necessary, it must be placed after the gravity thickeners.

The use of other thickening methods such as dissolved air flotation, basket or solid bowl centrifugation, low pressure belt filtration and the rotary drum system has increased because these methods can also give reliable and effective results when thickening WAS.

3.3.3 Particle Surface Charge and Hydration

Sludge particles have a negative surface charge and try to repel each other as they are brought together. Additionally, sludge particles weakly attract water molecules to their surface (hydration) either by weak chemical bonding or by capillary action. Although the water is only weakly held at the particle surface, it does resist thickening and interfere with dewatering.

Chemical conditioning is used to overcome the effects of surface charge and surface hydration. Typical chemicals are organic polymers; lime, ferric chloride and other metallic salts. Generally they act by reducing or eliminating the repulsive force, thus permitting the particles to come together or flocculate. Water can be more readily removed at a higher rate during the subsequent mechanical dewatering. Sludge conditioning is discussed in detail in Chapter 5.

3.3.4 Particle Size

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

- Greater repulsion between particles due to the larger area of negatively charged surface
- Greater attraction of water to the particle surface due to more sites for chemical joining.

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

Individual particles of activated sludge are usually aggregated to an extent through bio-flocculation. Table 3-5(5) shows the relative difficulty of removing water from an unflocculated primary digested sludge containing various particle size fractions. As can be seen, the Specific Resistance to filtration of the unfractionated sludge is dominated by the Specific Resistance of material under 5 microns in size, even though this material constitutes only about 14 percent by weight of the total solids. Specific Resistance is, in effect, a measure of the relative dewaterability of a sludge. The lower the Specific Resistance, the greater the sludge's dewaterability. Specific Resistance has been defined as the pressure required to produce a unit rate of flow though a cake having a unit weight of dry solids per unit area when the viscosity of the liquids is unity. Specific values are determined from laboratory filtration experiments.

Table 3-5. Sludge Dewatering as a Function of Particle Size

Mean Diameter	Specific Resistance	Percent of Total Particles
microns	sec^2/g	
Original, unfractionated sample	10.4×10^9	
> 100	2.3×10^9	10.2
5 - 100	4.6×10^9	75.5
1 - 5	13.8×10^9	8.5
< 1	-	5.9

Table 3-6(5) contains typical Specific Resistance values for different types of sludges, both chemically treated and untreated. Since the maximum Specific Resistance for feasible mechanical dewatering is normally quoted at $\leq 10.0 \times 10^7$ sec^2/g, none of these sludges would be readily dewaterable. Table 3-6 shows that Specific Resistance values can vary significantly. Experience indicates that properly conditioned raw primary sludge is almost always the most readily dewatered, followed by well-conditioned digested primary sludge and then activated sludges, in increasing order of difficulty.

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

Table 3-6. Specific Resistance of Various Types of Sludges

Type of Sludge	Specific Resistance
	sec^2/g
Raw	$10 - 30 \times 10^9$
Raw (coagulated)	$3 - 10 \times 10^7$
Digested	$3 - 30 \times 10^9$
Digested (coagulat...	$2 - 20 \times 10^7$
Activated	$4 - 12 \times 10^9$

3.3.5 Compressibility

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

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

3.3.6 Sludge Temperature

As sludge temperature increases, the viscosity of the water present in the sludge mass decreases. Viscosity is particularly important in centrifuge dewatering since sedimentation is the key component of the process (1). See Section 7.3 for further discussion.

3.3.7 Ratio of Volatile Solids to Fixed Solids

Sludges tend to dewater better as the percentage of fixed solids increases. One high-G centrifuge manufacturer uses the percentage of fixed solids as a key parameter in sizing equipment (R.T. Moll, Sharples-Stokes Div., Pennwalt Corp., personal communication, 1982). (See Section 7.3 for a description of low-G and high-G centrifuges.) According to this manufacturer, the sludge cake from centrifugal dewatering of an anaerobically digested mixture of primary and waste activated sludge shows a positive change of 5 percent in its solids concentration when the percentage of volatile solids in it decreases from 70 percent to 50 percent. However, since digestion also produces smaller particles, the higher surface area results in more moisture. The above approximation of volatile content to cake solids must be cautiously employed and should be pretested whenever possible.

3.3.8 Sludge pH

Sludge pH affects the surface charge on sludge particles. Hence pH will influence the type of polymer to be used for conditioning. Generally anionic polymers are most useful when the sludge is lime conditioned and has a high pH, while cationic polymers are most suitable at a pH slightly above or below neutral. In some cases, cationic polymers can be effective up to pH 12 and has been employed for lime stabilized sludge at New Haven, CT. Polymer technology is continuing to advance.

3.3.9 Septicity

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

3.4 Recirculation from Solids Processing

The return flows emanating from sludge thickening, digestion, conditioning, and dewatering will recycle TSS and BOD_5. If the primary clarification is not hydraulically overloaded, the majority of these solids will resettle in the primary clarifier. Contrary to popular opinion, there will be no significant increase in primary clarifier effluent TSS due to recycle loads of 25 percent or more of the influent TSS concentrations. The treatment plants at York, PA, Dubuque, IA, and New Haven, CT all experience TSS recycles up to and exceeding 100 percent of the influent TSS without impairing primary effluent TSS. However, if the primary clarifiers were hydraulically overloaded and/or allowed to go septic, high recycle levels of BOD_5 and TSS would be a serious problem.

Thus, the bulk of TSS in the recycle stream will resettle and can be contained. This is not true for the soluble BOD_5 fraction of the recycle. This recycle loading must be added to the anticipated primary effluent BOD_5.

Return streams should always pass through the primary clarifiers (if present) and through pre-aeration if it exists. Pre-aeration will enhance the removal of solids while freshening the wastewater.

3.5 References

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Chapter 4
Process Selection

4.1 Introduction

Above all, the design engineer must ensure that capacity limitations in the sludge processing system are not the direct cause of impaired effluent quality. That is, the design should provide for sufficient standby capacity or an alternative mode of sludge handling, whereby solids can be removed from the wet-end processing in an orderly manner -- even if the primary means of sludge disposal is unavailable or has failed in some manner. This criterion applies equally well to plants small and large, whether utilizing mechanical or non-mechanical means of sludge disposal.

Alternative methods (standby capacity) for non-mechanical methods of sludge dewatering can include:

- *Multiple units* - one standby unit is the best alternative. For example: two operating sand beds and one bed unit available as a spare.

- *Liquid land spreading* - land spread sites must be approved in advance and liquid storage must be available and consistent with state regulatory requirements. Sixty to 120 days storage may be required in areas where the ground becomes frozen.

- *Storage on-site* - per state regulatory requirements.

- Liquid haulage to another plant - procedure should be approved in advance by the alternate facility through written agreement between the two authorities.

Alternative methods for mechanical dewatering equipment can include:

- *Duplication of capacity or one standby unit* - recommended as best alternative.

- *Maintenance of existing pre-expansion dewatering equipment in addition to new equipment* - often this appears to be a feasible approach, but in reality fails since disuse results in disrepair and, eventually, the old equipment is inoperable when needed.

- *Liquid haulage* - satisfactory for small plants, provided the alternative mode has been guaranteed by permits and prior agreements and that it is suitable for year-round disposal.

- *Storage* - has limitations, except in smaller plants (the storage capacity must exceed the time required to repair equipment).

- *Contract disposal services* - not available in all areas.

The evaluation of sludge dewatering alternatives must consider the possibility of upstream and downstream processes being out-of-service. Consider, for example, a gravity thickener, which concentrates the raw primary sludge (RPS) and waste activated sludge (WAS) to 5 percent TSS, being out-of-service. If there is only one thickening unit, what alternative provisions are available to partially concentrate the sludge? Can downstream process units handle the more dilute sludge volume?

If there are two gravity thickeners, the off-line unit could reduce the underflow concentration to about 3.5-3.8 percent TSS due to the doubling of the unit loading ($kg/m^2/d$). Or, if there is a single combustion unit, is there an acceptable alternative method of raw sludge disposal available? If not, provision for lime stabilization should be included in the dewatering plan.

There is seldom a defendable reason for having only one dewatering unit in a wastewater treatment system, except in very small systems which have adequate sludge storage/disposal capacity by an approved alternative method. In no case should the lack of sludge dewatering equipment require storing sludges in the wet processing operations; this includes the sludge thickening units. Sludge concentration processes prior to dewatering should be duplicated unless it can be demonstrated that the dewatering equipment can process the more dilute, higher volume sludge in the operating time available. The unconcentrated sludge may reduce the dewatering unit's operating capacity by 50-70 percent.

4.2 Sludge Processing Methods/ Selection Procedures

When either evaluating or selecting a dewatering process, the designer must consider both prior treatment processes and subsequent disposal practices. A dewatering process cannot be evaluated without also examining the other processes involved in the overall solids handling system. Such an evaluation can be a complex procedure because of the vast number of combinations of unit processes available for thickening, stabilization, conditioning, dewatering, and ultimate disposal. Figure 4-1 presents a general schematic of a typical solids handling system and the unit processes most commonly used to perform each of these functions.

The strategy for selection of a dewatering process at either new or existing plants requires up to five stages of analysis, as shown in Figure 4-2. These stages represent a screening procedure in which the dewatering processes under consideration are given increasing scrutiny as more detailed cost, operational, and design data are collected and evaluated. The components of each of these stages are discussed below:

Stage 1 - Initial Screening of Dewatering Processes

First, a large number of factors should be reviewed so that incompatible processes can be eliminated prior to the initial cost analysis. Factors to be considered in the initial screening include:

- Compatibility with plant size and existing facilities
- Type and quantities of sludge produced
- Compatibility with the ultimate disposal technique (Selected dewatering process must be able to produce required cake solids concentration)
- Compatibility with available labor and land
- Degree of conditioning required
- Environmental considerations
- Field experience with processes at other similar operating installations.

Stage 2 - Initial Cost Evaluation

Based on the best estimates of design and operational criteria for the feasible dewatering processes, an initial cost evaluation should be conducted. In some cases, 10 to 20 complete solids handling alternatives, which may include 4 to 5 different dewatering processes, are evaluated in this initial stage. In general, no more than 3 to 5 of the lowest cost alternatives are selected for more detailed evaluation.

Stage 3 - Laboratory Testing

Laboratory testing may be conducted on the dewatering processes selected in Stage 2 to refine design criteria for the more favorable dewatering techniques. This laboratory testing may be conducted at the plant or by equipment manufacturers in their laboratories. This testing will have limited value unless conducted on a representative sludge that has not undergone change in transport. Laboratory tests should be conducted near the source of sludge in order for the full range of fresh sludge characteristics to be tested. A test on one sample of sludge has negligible value and could be a cause of design errors. Substantial day-to-day variation in mixed sludge characteristics can occur. "Too good to be true" results most often are just that. (See Chapter 5 for more information on laboratory tests.) Field testing is preferred except for smaller plants.

Figure 4-1. General schematic for solids handling, showing most commonly used methods of treatment and disposal.

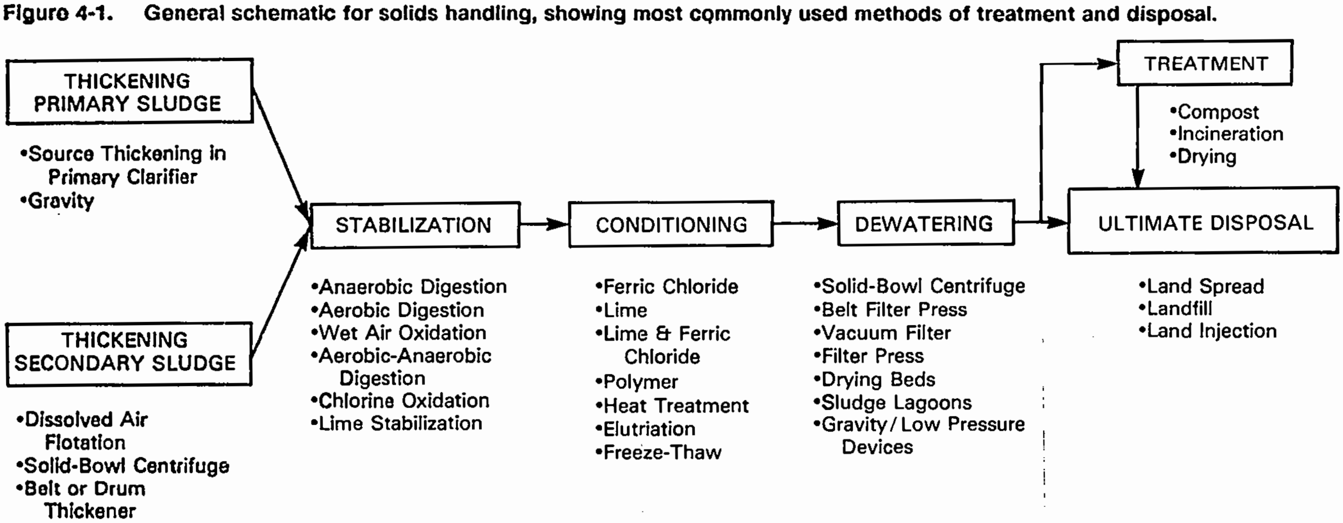
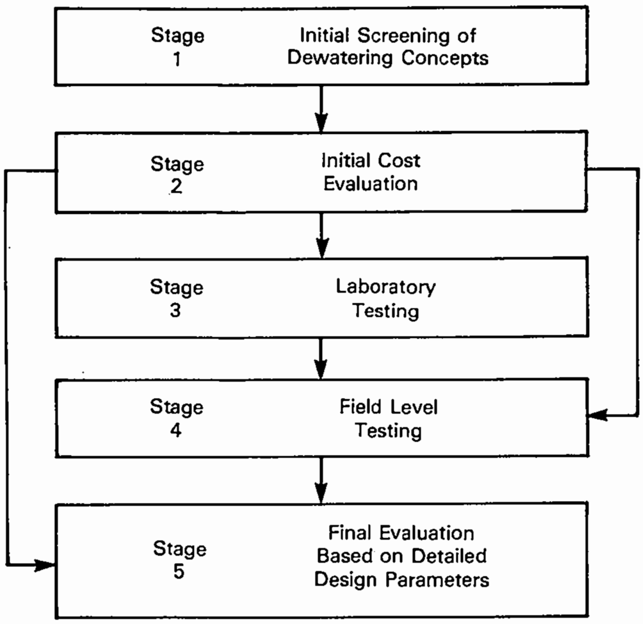


Figure 4-2. Five stages of analysis in selection of a dewatering process.



Stage 4 - Field Pilot Testing

If the plant is large and/or the cake moisture content is critical (and more than one dewatering method may be feasible), pilot studies are often warranted. Since the sludge dewatering properties of even apparently similar sludges may vary widely, pilot studies greatly reduce the risk of improperly selecting and sizing dewatering equipment. The cost of a thorough evaluation is small compared to the benefits gained.

If it is necessary to test two types or more of comparable dewatering equipment, the tests should be conducted simultaneously to eliminate potential differences related to the sludge composition. Sludge variations, due to a number of reasons, can distort the comparison. An example, shown in Table 4-1, is the result of two series of tests conducted at the same plant in Ohio.

Table 4-1 indicates that two dewatering studies at the same facility produced widely different results. Moreover, there was not a similar comparative difference between the two types of equipment for the winter and summer testing. The differences found must also be considered in light of the degree of optimization achieved. Short-term testing may not have fully evaluated the range of operation or optimized the critical chemical conditioning step.

The centrifuge data in Table 4-1, which was produced by full-scale operation, indicate the magnitude of the problem that could have been encountered had the centrifuge installation been sized on winter test performance results. Production rate and cake solids content were much lower in the summer tests. The differences were a result of storm flows adding inert material to the sludge and changes in industrial discharges. In the winter tests (Series I), storm flows had added inerts to the sludge; in the summer tests (Series II), a high TSS discharge from a brewery had a more adverse impact on the performance of the centrifuge than on the performance of the diaphragm plate press.

Ideally, pilot testing should be carried out over an extended period of time. However, extended testing is often not practical. Test programs should evaluate a sufficiently wide range of PS:WAS ratios to ensure testing of worst-case situations, preferably during colder weather when the sludge water viscosity and secondary sludge yields are higher. Further, a full range in operating capacity should be investigated to determine the effect on cake solids and capital and operating costs.

Stage 5 - Final Evaluation Based on Detailed Design Parameters

After Stage 4 is completed, accurate scale-up and sizing of equipment is performed by the design engineer with the aid of the equipment manufacturer. At this time, estimates of the capital cost, labor, energy, chemical, and maintenance material requirements for the dewatering process under consideration can be refined. This information can be supplemented with data from other plants using the same process. The researching of similar equipment performance and the manufacturers' service record is highly recommended. Additionally, the operating utility can make input from performance and operational problems experienced in Stage 4 field evaluations. Based on accurate capital and operation and maintenance cost information, a final cost evaluation can be made in conjunction with an evaluation of other parameters. Stage 5 concludes with selection of the dewatering process and, in many cases, the preferred manufacturer. All generic equipment is not created equally.

The equipment and supplier selected should have widely demonstrated the capability to meet the design requirements in either similar plants or by adequately supervised pilot studies at the subject facility. When a new design of equipment is employed, the manufacturers' prior practices need to be carefully scrutinized and adequate safeguards provided to the utility. Evaluations properly conducted will not stifle new developments.

Throughout this five-stage process, many trade-offs will have to be made. In many cases, the total annual cost of two or more solids treatment systems are essentially identical (±10%), and the decision must be made on some basis other than cost. Frequently, such a decision is based upon capital vs.

Table 4-1. Comparative Dewatering Results for Two Test Periods

	PS:WAS Ratio	Feed Rate		Cake TS	SS Recovery	Polymer Cost	Total Chemical Cost
		l/s	kg TS/hr	percent	percent	$/Mg	$/Mg
Series I - Winter							
C-1	1:1	3.8	314	24.3	92	31.03	31.03
C-2	1:1	5.0	382	23.6	86	23.27	23.27
C-3	1:1	6.3	477	17.3	85	19.64	19.64
DPP-1[1]	1:1	65.0[2]	20.9[3]	21.9	99+	11.64	28.31
DPP-2[1]	1:1	80.0[2]	14.2	27.1	99+	13.18	29.85
Series II - Summer							
C-1	1:1	2.5	282	17.6	97	17.09	17.09
C-2	1:1	3.8	423	13.3	96	29.17	29.17
DPP-1[1]	1:1	-	15.1[3]	24.3	99+	5.81	22.48
DPP-2[1]	1:1	-	19.5	27.2	99+	8.64	25.31

C = 0.74 m diameter x 2.34 m long solid bowl centrifuge.
DPP = Lab diaphragm plate press.
Polymer addition w/precoat @ 12% of sludge solids.
$FeCl_3$ @ $0.25/kg
Precoat @ $0.66/kg
[1] Used 3-4% $FeCl_3$ to improve flocculation.
[2] Press time, min.
[3] Cake discharge, kg TS/m^2 of plate area/hr.

O&M cost considerations, ease of equipment operation, energy requirements, performance, or other factors such as prior plant experience with similar equipment. A point to keep in mind is that the decision is often not clear-cut.

The overall complexity of analysis will vary depending on the size of the plant and whether a new solids handling system is being designed or an old one upgraded. If the solids handling system is completely new, there will probably be fewer constraints on the processes to be evaluated, conditioning method to be used, and ultimate disposal techniques to be considered. In other situations, if the entire treatment facility is new, or if it is being upgraded from primary to secondary, sludge of the correct composition will not be available to conduct field tests. Stage 4 is generally not conducted for most of the small capacity plants, those less than 0.13 m^3/s (3 mgd). For the small plant, it is usually more economical to design facilities based on laboratory or bench-scale testing (often performed by the manufacturer of the equipment) and after experiences at other plants, and using generous factors of safety in design, than it is to conduct the field-scale testing. The field-scale testing may result in a recommendation for smaller and thus less expensive equipment, but the reduction in cost is unlikely to offset the extra time and cost of the full-scale testing.

While past experiences with similar equipment and facilities can provide useful input, engineers must ensure that they are current in the process and equipment technology under consideration and that design requirements are similar. A design practice of using the same process technology for all plants is highly questionable. It is the design engineer's responsibility to make independent evaluations of the equipment's performance and not to accept other evaluations by parties with a financial interest in the outcome without first checking the reduced data.

All designs for sludge processing and dewatering systems should undergo a thorough "what if" evaluation. That is, evaluation of all probable situations that could occur in the future plant operation. The limiting conditions should be defined and analyzed for any adverse impact on liquid processing capabilities. Sludge storage in the liquid treatment process is not an acceptable alternative, since effluent quality degradation will occur soon thereafter. Further, the hydraulic peaking capacity of the clarification operations will be reduced by sludge storage. Sludge processing equipment should be selected and sized to process the sludge produced in a timely manner.

A formal ranking of the alternative dewatering methods versus key selection criteria is recommended. The key criteria should have appropriately weighted values. The formal ranking results should be internally and sometimes independently critiqued to ensure all considerations have been adequately evaluated.

4.3 Operational Selection Criteria

The criteria employed for selecting a sludge dewatering process are complex and will vary from site to site and with the size of the plant. Given similar circumstances, engineers will often select different dewatering processes based on their past experiences and personal preferences. Often there is more than one correct process selection, or the correct selection could only be established by exhaustive testing and engineering economic analysis.

The selection criteria for smaller plants are considerably different than those for larger facilities. In many cases, other considerations, such as transport and land availability, may have a more significant impact than economics. Table 4-2 presents various selection criteria versus plant size. However, site-specific conditions may change the relative importance of some of the criteria. For example, a drying bed in a highly developed area would be more objectionable than it would be in a semirural setting or in a heavily industrialized area.

Table 4-2. Operational Selection Criteria for Sludge Dewatering Processes

Plant Size	Key Criteria
Small < 0.08 m^3/s (<2 mgd)	Minimum Mechanical Complexity Local Repairs and Parts Minimum Operator Attendance Reliable Without Skilled Service Unaffected by Climatic Factors Large Excess Capacity Handleable Cake
Medium 0.08 - 0.44 m^3/s (2 - 10 mgd)	Low Operator Attendance Local Repair and Parts Transportable Cake Without Nuisance Mechanical Reliability Competitive O&M Costs Drier Cake
Large >0.44 m^3/s (>10 mgd)	Lowest O&M Costs/ton Dry Solids Lowest Capital Costs/ton Dry Solids Driest Cake High Output/Unit Mechanical Reliability Transportable Cake Without Nuisance

General Considerations

- Compatibility with existing equipment with long-term sludge disposal
- Long-term serviceability/utility
- Acceptable environmental factors
- Good experience at other operating installations
- Competence and quality of local operator and service personnel
- Compatibility with plant size
- Acceptance by user and regulatory agency
- Availability and need of manufacturer's services

The classification of plants as small [<2 mgd (0.08 m^3/s)], medium [2-10 mgd (0.08-0.44 m^3/s)] and large (>10 mgd (0.44 m^3/s)] is rather arbitrary and is used merely as a generalization to segment applicable technology. The personnel at a 0.11-m^3/s (2.5-mgd) advanced wastewater treatment plant treating wastewaters to 5 mg/l BOD_5, 5 mg/l TSS, 1 mg/l NH_4N, and 1 mg/l total phosphorus may be much more qualified to operate and maintain a mechanical dewatering unit than the personnel at a 0.53-m^3/s (12-mgd) plant using trickling filters to meet 30/30 criteria. The anticipated quality and quantity of the O&M staffing are important criteria for the engineer to consider in the selection of process technology.

In smaller plants, the lowest initial cost may not be the best selection if it requires continuous operator attendance, is mechanically complex, and cannot be repaired locally. The impact of these considerations can easily offset the advantage of an alternative lower cost dewatering process. A mechanical dewatering process that is operating at 200 kg dry solids/hr, but requires continuous operator attention, could result in over $50/ton ($55/Mg) operating labor costs plus the maintenance costs.

4.3.1 Compatibility With Existing Facilities

Existing facilities, which must be considered in evaluating dewatering processes, include:

- Type of dewatering equipment presently used, its useful remaining life, and its compatibility with future requirements
- Existing conditioning, chemical storage and feed facilities
- Existing building used for dewatering and ancillary equipment
- Existing site constraints
- Existing sludge transport facilities.

4.3.1.1 Existing Dewatering Equipment

Existing dewatering equipment customarily plays a major role in the selection of additional equipment, particularly if space has been provided for expansion of the present dewatering facilities. If existing equipment is providing satisfactory performance (from both a cost and operational standpoint), and if the product cake is suitable for the ultimate disposal technique, in all likelihood the same dewatering process would be appropriate for the expansion plan. This would be particularly true if the dewatering facilities had been designed to accommodate more equipment of the same type. In perhaps the majority of dewatering operations, existing equipment is performing unsatisfactorily and requires more chemicals or energy than originally anticipated. In other cases, the sludge characteristics have adversely changed, and the existing equipment cannot be operated at the original design capacity. In some cases, existing equipment cannot perform as well or as efficiently as some of the newer but similar equipment available, or the cake produced by existing equipment is not suitable for the future ultimate disposal technique.

In a large percentage of the expansions of dewatering facilities, the plant staff is dissatisfied with the operation of the existing equipment. Typical situations are: (1) vacuum filter installations where lime coating of the filter media, filter drum, and filtrate piping presents an expensive and continuing maintenance problem; (2) filter press installations that often have chemical requirements substantially higher than originally expected; (3) older existing solid-bowl centrifuge installations where a great deal of scroll maintenance is required due to abrasive wear and/or where the operating performance is poor; and (4) drying beds or lagoons where odors, negative visual impact, intensive labor requirements, or difficulty with sludge removal make the process an operations problem for the plant staff. These types of problems can cause headaches for the operation and maintenance staff and can decrease effective dewatering capacity. Also, operating costs increase when equipment must be taken out of service for repairs and/or cleaning.

Variation in sludge characteristics after design and installation of equipment, and therefore variation in the ability of the sludge to be dewatered, presents a particularly vexing problem. Very often, variation of sludge characteristics leads to higher conditioning and energy requirements than originally projected, and, in some cases, the inability to produce a dewatered cake suitable for ultimate disposal. Sometimes equipment must be operated at less than design capacity due to changed sludge characteristics. An evaluation should be conducted to determine the likelihood and severity of changes in sludge feed rate and characteristics. If significant variations are anticipated, equipment, such as the centrifuge, that is less sensitive to such changes should be selected.

More often than not, equipment technology will have advanced since the original equipment was installed. In this case and particularly where there is a high degree of owner/operator dissatisfaction, the obsolete equipment should be removed and replaced by modern equipment. Where selection evaluation indicates that there is an acceptable alternative to the original, unsatisfactory process technology, it is generally advisable to employ a different process for dewatering to ensure plant cooperation. Where possible, extensive test trials to demonstrate the advanced technology to the plant operators are recommended.

<u>Obsolete equipment does not provide good standby capability!</u> There is a natural resistance to upgrading or maintaining old, obsolete equipment. If there is significant operator resistance to use of obsolete equipment, it will not be in a serviceable condition when needed. When faced with obsolete equipment, the engineer and the city should "bite the bullet" and provide a totally new dewatering station.

4.3.1.2 Existing Building Used for Dewatering Equipment

If the original design allocated space for an anticipated expansion, there are generally few problems associated with installing newer equipment of similar size. Newer equipment, built to handle a given sludge volume, is usually lighter than similar older equipment. However, if larger equipment is being installed, then the suitability of the housing facility will be a primary concern.

The engineer must consider the present building's structural capacity for replacement equipment, and whether the building has sufficient headroom and working space for the equipment being considered. Dewatering equipment, such as solid bowl centrifuges, belt filter presses, and filter presses frequently discharge dewatered solids downward. These machines can be incompatible with buildings with low roofs, because in some cases they require elevated mounting to provide space for conveyor belts under the equipment.

Further, heavy equipment, such as a filter press, may not be compatible with a building originally designed for a centrifuge installation, even though both have bottom discharge of cake solids. Centrifuges may require greater structural support than belt presses. When there is an existing overhead crane, the new equipment may exceed the allowable crane capacity, and it would also need to be replaced. Lack of operator working space is a means of ensuring poor maintenance and attendance. Crammed facilities should be avoided at the cost of new or expanded building space.

4.3.1.3 Existing Site and Environmental Constraints

Drying beds and sludge lagoons require considerable land area. Expanded use of these processes may not be practical if land is unavailable, or if environmental constraints make continued use unacceptable. In some cases, existing beds or lagoons can be used in conjunction with a different dewatering process.

For example, drying beds could be used to produce a dry product in the warm, drier periods of the year and the mechanical equipment would be used when the weather is adverse to dewatering sludge on the drying beds. The drying beds would constitute the backup in the winter, and vice versa.

Design engineers need to ask the city and themselves several questions before deciding on a dual-technology alternative. Typical questions could be:

1. Which method requires the least labor? Is this important?

2. Is there adequate staff at the plant now? Will there be?

3. Are the drying beds well maintained now? If not, is there any reason maintenance would improve if there were an alternate mechanical unit available?

4. Is the absence of chemicals a sufficient driving force to result in the use of beds in warmer weather? Does the additional labor offset the cost of chemicals and convenience of a mechanical dewatering station?

5. Is there an advocate or user who wants the drier sand bed solids as opposed to a wetter cake?

Unless such questions are posed and truthfully answered, most dual-technology dewatering technology systems are not fully used and one system falls into disuse. If the design engineer cannot justify the money to refurbish the older facilities, they should be replaced.

The local environmental setting may force the closing down of a good operating system such as a sand bed. An alternative may be to enclose the sand bed and ensure that it is properly ventilated and that odor control is provided. Enclosure also brings additional capacity to the same system since weather impacts are minimized and the sand beds could be heated. Further, increase in capacity can be achieved with small dosages of polymer.

4.3.2 Process Compatibility With Size of Plant

Use of uncomplicated sludge handling systems increases the chances for successful operation in any size of plant. Complex equipment is especially unsuited to small plants for several reasons. First, the amount of operator time available generally decreases as plant size decreases. Second, small plants may not have operations and maintenance personnel with the required skills. Third, less complicated equipment is generally less expensive to purchase. Since standby capacity of about 100 percent is often provided for small plants (usually with duplicate units rather than employing a single large unit), it is more economical to choose the less complicated, less expensive system.

The foregoing discussion presents a basic approach to selection of the dewatering process. Generalized guidance based on results at plants across the United States is summarized in Table 4-3, which presents compatibility of different dewatering techniques with various plant sizes. Designers should use the information presented in Table 4-3 only as a guide. Every plant must be considered independently, since site-specific considerations can have a large influence on the dewatering process. For example, drying beds and sludge lagoons may be cost-effective at a plant larger than 0.44 m^3/s (10 mgd), if weather is favorable and land available. Using a structured approach to evaluate the alternatives, in conjunction with the costs of the ultimate solids disposal method, will sort out the appropriate technology.

Table 4-3. Compatibility of Dewatering Process with Plant Size

	Plant Size, m^3/s		
	<0.04	0.04-0.44	>0.44
Belt and Drum Thickeners	X[1]	X	X
Solid Bowl Centrifuge		X	X
Belt Filter Press	X[2]	X	X
Vacuum Filter		X	X
Filter Press		X	X
Drying Beds	X	X	
Sludge Lagoons	X	X	

[1] Suitable for land spreading or injection of sludge.
[2] Only low pressure press is commonly used in this flow range.

4.3.3 Process Compatibility with the Ultimate Disposal Technology

The ultimate disposal method of the residual solids often dictates the process selection. The most common methods of disposing of the cake are shown in Table 4-4. The end-product of each dewatering method is rated in terms of the product acceptability vis-a-vis the ultimate disposal of the residuals. However, this table is not meant to establish the compatibility of the dewatering process and the end-disposal means. For example, combustion of a sand bed product of 30-40 percent TS is very economical, but the use of sand beds to dewater sludge solids for combustion may be impractical. On the other hand, a compost operation may be better equipped to process the periodic and somewhat variable sand bed product which can be stockpiled until needed.

While all sludge cakes meeting EPA's stability regulations (see 40CFR257) can be land spread, drier solids will be easier and less costly to transport and apply to the land. The drier the cake, the less the probability of any nuisances such as odors, insects, or liquid runoff. Ideally, sludge applied to land should either be sufficiently dry so that the spreader can break it into small pieces or should be a liquid that can be evenly spread. Injection of liquid sludge can eliminate nuisances, but it also can be limited by weather conditions. If the land is to be "worked" after the sludge application, then the cake solids concentration is less of a concern, except where the sludge cake must be stored for part of the year. In this case, a drier cake is preferred.

Table 4-4. **Suitability of Cake Produced by Dewatering Processes for Various Ultimate Disposal Options**

Dewatering Process	Cake Solids %	Land Spread	Land Spread Inject	Landfill	Combustion/ Drying	Compost
Lagoon	15 - 40	1	-	2	3	2
Sand Beds	30 - 60	1	-	1	1	1
Vacuum Assisted Beds	10 -	1	-	3	3	3
	16	2	-	21	3	2
Screw Press	12 -	1	-	3	3	3
	20	2	-	2	3	2
Paved Beds	30 - 60	1	-	1	1	1
Belt Press	18 -	2	-	2	3	2
	24	2	-	2	2	2
Centrifuge	18 -	2	-	2	3	2
	24	2	-	2	2	2
	5-7	1	1	3	3	3
Vacuum Filter	16 - 20	2	-	2	3	2
Filter Press	26 -	1	-	1	2	2
	34	1	-	1	1	1
Drum/Belt Thickeners	5 - 8	1	1	3	3	3

Key: 1. Good
2. Satisfactory
3. Inappropriate

Note: The ratings apply to the product concentration and not the methodology employed to achieve the cake solids.

Landfill operations are becoming more selective in the materials accepted. Wet cakes, which may result in liquid discharges, are particularly discouraged. A minimum cake of 18 percent TS is considered acceptable for a narrow trench landfill. However, higher percent TS may be required in some areas of the United States, and in the future higher percent TS may be required in many areas. West Germany currently requires 35-40 percent TS for landfill.

Dewatering of sludge for combustion or drying requires that the solids content be equal to 24 percent TS, preferably 28-30 percent for an economical operation. On a typical raw primary sludge (60 percent) and WAS (40 percent) mixture, only drying beds and filter presses can reliably deliver 24 percent TS or drier cakes. While many multiple hearth furnaces (MHF) in the United States can burn 24-25 percent TS sludge cakes autogenously, the stack outlet gases are below 480°C (900°F). Some new state regulations now require a 650-760°C (1,200-1,400°F) stack outlet temperature to ensure deodorization. Fluid bed reactors (FBR) require about 28 percent TS with heat exchangers to preheat the air and 34 percent TS without heat exchangers to burn raw primary and waste activated sludge autogenously. Combustion gases from FBR units always exceed 800°C (1,470°F) due to process configuration.

The combustion capacity of multiple hearth furnaces and fluid bed reactors as a function of sludge cake solids concentration is shown in Figure 4-3. Typical excess air for MHF is 75-125 percent and 30-40 percent for FBR. The moisture content also has a direct impact on the size of the furnace. A 20 percent TS cake (4 kg H_2O:kg TS) will require a 56 percent larger furnace than a 28 percent TS cake (2.57 kg H_2O:kg TS) for the same dry solids capacity.

Typical heat content of a 60:40 to 50:50 mixture of PS:WAS is 5,555 $\pm$ 278 kcal/kg VSS (10,000 Btu/lb $\pm$ 500 Btu/lb) and the VSS/TSS ratio is usually 72 $\pm$ 5 percent for raw sludges. A more precise means of expressing the effect of cake solids content on an exhaust temperature is to use the wet cake heat content, kcal/kg (Btu/lb) wet cake, and the required excess air and other losses. The relationship of kcal/kg wet cake, percent TS, and exhaust temperature for autogenous conditions are shown in Figure 4-4.

An outlet temperature of 540°C (1,000°F) for a FBR is the equivalent of preheating air to about 540°C (1,000°F) with a 815°C (1,500°F) stack outlet temperature before the heat exchanger. The autogenous conditions noted in Figure 4-6 are based on 50 percent excess air. FBRs will employ 30-50 percent excess air.

The effect of excess air on heat losses in a combustion unit is illustrated in Figure 4-5. For example, if a furnace is operating at 540°C (1,000°F) outlet gases, the heat loss at 40 percent excess air is about 6 percent of the input. At 125 percent excess air, the loss is 300 percent higher or 18 percent of the input.

The effect of operating exhaust temperature and excess air on fuel consumption is quite apparent in

Figure 4-3. Multiple hearth furnace and fluid bed reactor capacities.

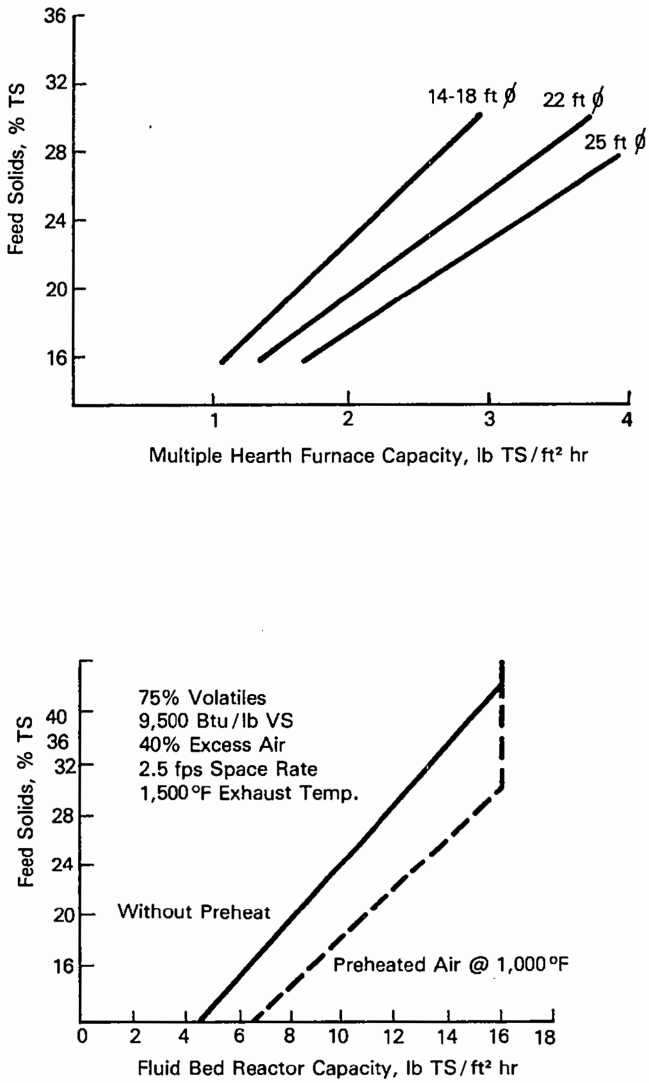
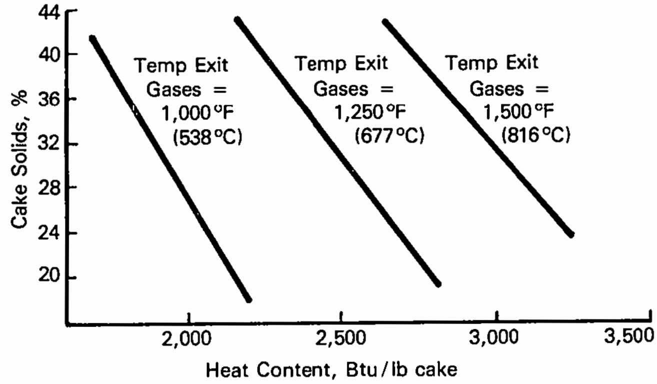


Figure 4-4. Conditions for zero fuel.[a]



[a]Assumed conditions:
1. Excess air = 50%
2. Misc. losses = 5%

Figure 4-5. Heat loss due to excess air.

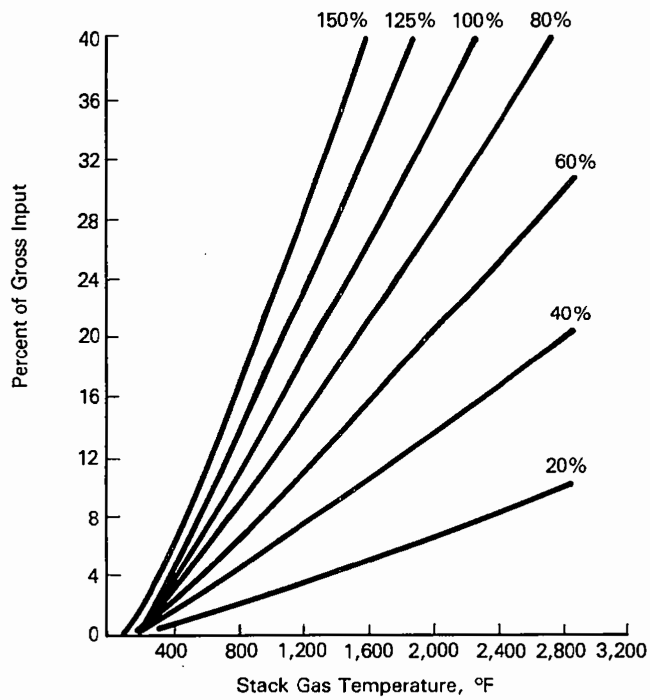


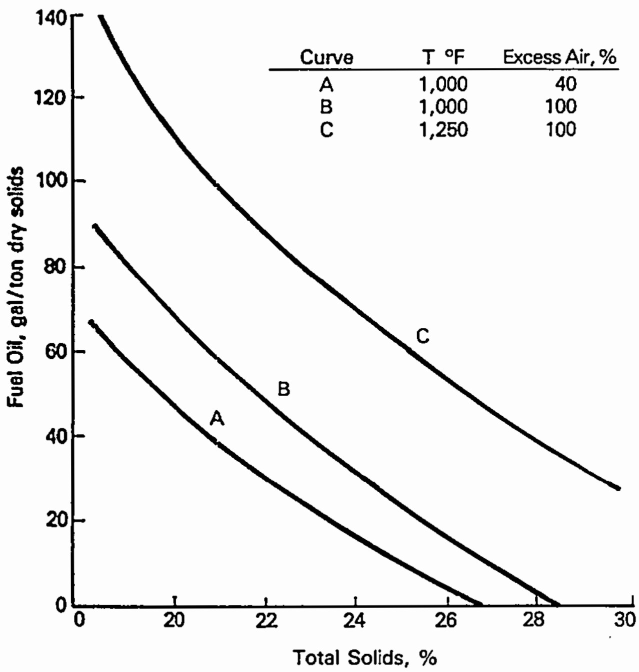
Figure 4-6. Clearly, if 540°C (1,000°F) at 100 percent or more excess air is necessary, the sludge should be dewatered to >26 percent TS to reduce fuel requirements to zero. Curves B and C represent fuel requirements for MHF @ 540°C (1,000 °F) and 680°C (1,250°F) while Curve A is for a FBR @ 815°C (1,500°F) using a preheater. Table 4-5 presents the dewatering and combustion costs for an array of different dewatering systems, which illustrates the impact of capital and operating costs on the overall cost of thermal sludge disposal.

Compost operations can sometimes suffer even more than combustion from wet cakes. This is particularly true in colder, wetter regions and where the compost piles are not protected from the elements. Since it is important to maintain the compost in a specific moisture and temperature range, it is necessary to recycle the drier product to bulk the sludge cake and to reduce the average moisture content. The wetter the cake, the higher the recycle or new bulking agent, the more the pile is cooled and the less economical the composting process.

An example of the effect of sludge cake moisture content on the recycle rate is demonstrated by Equation 4-1:

Minimum sludge + recycle TSS = 45% TS (4-1)
Sludge cake concentration = 18% TSS & 30% TS
Sludge cake quantity = 10 Mg/d

Figure 4-6. Fuel consumption vs. stack temperature and excess air.



Recycle product concentration = 55% TS

Dry Solids Balance:
(Mass of
Mixture/% TS) = (Cake/% TS) + (Recycle/% TS)

@ 10% TS:

(X/0.45) = (10/0.18) + [(x - 10)/0.55]

X = 93 Mg/d dry recycle solids
or 169 Mg/d product recycle @ 55% TS

@ 30% TS:

X = 37.6 Mg/d dry product solids
or 68 Mg/d product recycle @ 55% TS

At some point, further dewatering may have very little benefit since the bulking requirements for air flow may govern. The recycle rates are also highly dependent on climatic conditions and the design of the compost process. Areas of high temperature and high evaporation rates will handle a wetter cake more economically because they can produce a drier product recycle. Enclosed compost systems in colder, wetter areas do not have prohibitive recycle rates because the recycle product moisture approaches the minimum feed moisture (sludge + recycle).

4.4 Sizing of the Dewatering Process

Where cake solids content is the governing criterion, operating at a rate below the rated capacity of the dewatering unit will increase the solids content. In some cases, a significant increase in solids content will result from only a 20-25 percent reduction in capacity. The rated capacity of manufacturers' units will generally be substantially higher than that which should be used for producing the driest cake.

When cake solids are not critical, the polymer cost to maintain 90-95 percent TSS recovery will be the governing criterion. The ability to readily move from maximum dry solids content to maximum solids recovery rate will vary from machine to machine. While the centrifuge is flexible in this regard, vacuum filter and belt and filter press operations must lie in the range of cake dischargeability. Higher polymer dosages can increase belt press capacity although this may reduce cake solids. Filter press operation may be quite inflexible in this regard due to cake dischargeability criteria.

Table 4-5. Dewatering-Combustion Costs - Excluding Operating Labor (7,500 dry tons/yr @ 60:40 PS:WAS)

Average Dewatering Costs/Ton Sewage Solids - 1986

Method	Chemical	Cake % TS	Cost, $/ton Maintenance	Power	Chemical	Total
Centrifuge/Belt Press	P	18	3.00	5.00	16.00	24.00
Centrifuge/Belt Press	P	22	3.00	4.00	16.00	23.00
Centrifuge/Belt Press	P+C	30	4.00	5.00	27.00	36.00
Diaphragm Plate Press	P	30	4.50	2.50	20.00	27.00
Recessed Plate Filter Press	F+L	32	4.50	2.75	30.00	37.25
Diaphragm Plate Press	P+A	45	5.00	3.00	20.00	28.00

Average Combustion Costs for FBR - 1986
(1,000°F - 50% Excess Air)

Method	Chemical	Cake % TS	Cost, $/ton Maintenance	Power	Fuel	Total
Centrifuge/Belt Press	P	18	7.50	5.00	74.80	87.30
Centrifuge/Belt Press	P	22	7.00	3.90	35.00	45.90
Centrifuge/Belt Press	P+C	30	6.50	3.10	0.00	8.60
Diaphragm Plate Press	P	30	6.50	2.70	0.00	9.20
Recessed Plate Filter Press	F+L	32	7.00	3.30	6.30	16.60
Diaphragm Plate Press	P+A	45	6.50	2.70	0.00	9.20

Combined Dewatering - Combustion Cost Summary

			Cost, $/ton				
			Dewatering		Combustion		
Method	Chemical	Cake % TS	Capital	Operating	Capital	Operating	Total
Centrifuge/Belt Press	P	18	6.00	24.00	33.60	87.30	150.90
Centrifuge/Belt Press	P	22	6.00	23.00	28.90	45.90	103.80
Centrifuge/Belt Press	P+C	30	7.80	36.00	25.20	8.60	77.60
Diaphragm Plate Press	P	30	9.00	27.00	24.60	9.20	69.80
Recessed Plate Filter Press	F+L	32	11.10	37.25	26.20	16.60	89.45
Diaphragm Plate Press	P+A	45	11.10	28.00	23.30	9.20	71.60

Operating Cost Basis:
Polymer (P) @ $2.00/lb
Ferric Chloride (F) @ $0.15/lb
Lime (L) @ $0.05/lb
Coal (C) @ $60.00/ton
Fuel @ $7.50/$10^6$ Btu
Sludge @ 9,500 Btu/lb VS
Coal @ 12,000 Btu/lb
Ash (A) - no cost

Capital Cost Basis:
Installed equipment, no buildings or foundations.

$/ton x 0.9072 = $/Mg
°C = (5/9) (°F-32)

Chapter 5
Conditioning

5.1 Introduction

Conditioning prior to dewatering involves the chemical and/or physical treatment of sludge to enhance water removal and improve solids capture. The three most common conditioning systems use inorganic chemicals, organic polymers, or heat. Table 5-1 shows and compares the effects of conditioning processes on a mixture of primary and waste activated sludges.

Conditioning always has an effect on the efficiency of the dewatering process that follows (1). Any evaluation of the conditioning process must therefore, take into consideration capital and operating and maintenance costs for the entire system. These costs include the impact of sidestreams on other plant processes, the plant effluent and resultant air quality.

Some treatment plants are required to remove phosphorus, although less of these plants must do so today than a few years ago. This is both because of state bans on the use of detergents containing phosphorus and the generally decreased use of phosphorus in household products. Phosphorus is often removed by the addition of chemicals, including ferric chloride, aluminum sulfate (alum), sodium aluminate, and lime with some kind of polyelectrolyte to facilitate coagulation and settling. Because of the enormous materials handling difficulties, the interest in using lime has declined.

Precipitating phosphorus can as much as double the amount of sludge requiring treatment and disposal. The amount of chemicals that must be added is a function of the amount of phosphorus needing removal (see Chapter 3). Fortunately, with the quantity of phosphorus in wastewater diminishing, this quantity is decreasing.

If precipitating phosphorus removal is contemplated, laboratory/pilot tests should be performed to determine the mass and volumes of sludge to be expected, the degree to which the sludge can be thickened, and how well the sludge can be conditioned and dewatered if appropriate. While in every instance there will be more sludge to contend with (an additional 30 to 100 percent), the sludges will thicken and dewater differently. Lime sludges readily thicken and dewater, while hydroxide sludges (ferric chloride, alum, sodium aluminate) usually thicken and dewater poorly - requiring considerable conditioning with polymers. In fact the best analogy is that difficult-to-handle hydroxide sludges behave like a poor quality activated sludge. It is not unusual also to find a hydroxide sludge that will at best dewater to a

Table 5-1. Effects of Conditioning with Inorganic Chemicals, Organic Polymers, or Heat on a Mixture of Primary and Waste Activated Sludge

	Inorganic Chemicals	Organic Polymers	Heat
Conditioning mechanism	Coagulation and flocculation	Coagulation and flocculation	Alters surface properties and ruptures biomass cells, releases chemicals, hydrolysis
Effect on allowable solids loading rates	Will increase	Will increase	Will significantly increase
Effect on supernatant stream	Will improve suspended solids capture	Will improve suspended solids capture	Will cause significant increases in color, suspended solids, soluble BOD, NH_3-N, and COD
Effect on manpower	Little effect	Little effect	Requires skilled operators and a strong preventive maintenance program
Effect on sludge mass	Significantly increases	None	Reduces present mass but may increase mass through recycle

final cake solids concentration of 10 to 15 percent on a centrifuge, belt press, and vacuum filter. These "still wet" sludges may be hard if not impossible to lift and unacceptable for landfilling.

5.2 Factors Affecting Conditioning

Wastewater solids are comprised of screenings, grit, scum, and sludges. Wastewater sludges consist of primary, secondary, and/or chemical solids with various organic and inorganic particles of mixed sizes. The sludges each have various internal water contents, degrees of hydration, and surface chemistry. Sludge characteristics that affect dewatering (and for which conditioning is employed) are particle size and distribution, surface charge, and particle interaction. Furthermore, such things as biopolymer production, degree of filamentous growth, primary:secondary sludge ratio, and inorganic content

Particle size is considered to be the single most important factor influencing sludge dewaterability (1). As the average particle size decreases, primarily from mixing or shear, the surface/volume ratio increases exponentially (1). Increased surface area means a greater hydration, higher chemical demand, and increased resistance to dewatering. Figure 5-1 shows relative particle sizes of common materials.

Raw municipal wastewater contains a significant quantity of colloids and fines that because of their size, 1 to 10 microns, will almost all escape capture in primary clarifiers if coagulation and flocculation are not employed. Secondary biological processes, in addition to removing BOD, also partially remove these colloids and fines from wastewater. As a result, biological sludges, especially waste activated sludges, are difficult to dewater and have a high demand for conditioning chemicals.

A primary objective of conditioning is to increase particle size by combining the small particles into larger aggregates. Since sludge particles are typically negatively charged and repel rather than attract one another, conditioning is used to neutralize the effects of this electrostatic repulsion so that the particles can collide and increase in size.

Conditioning is a two-step process consisting of coagulation and flocculation. Coagulation involves destabilization of the sludge particle by decreasing the magnitude of the repulsive electrostatic interactions between particles. This process occurs through compression of the electrical double layer surrounding each particle. Flocculation follows coagulation and is the agglomeration of colloidal and finely divided suspended matter by gentle mixing.

If the flocculated sludge is subjected to stress, floc shearing can occur. Therefore, mixing should provide just enough energy to disperse the conditioner throughout the sludge and bring the particles and colloidal suspensions together. Consideration should be given to providing individual conditioning for each dewatering unit, since it is neither always economical nor good practice to provide one common conditioning unit for several dewatering units. Problems can arise in balancing the flow rates of the various streams when starting up or shutting down individual units. The location of the conditioning unit, relative to each dewatering device, requires optimization.

The amount of conditioning required for sludges depends on the processing conditions to which the sludge has been subjected and on the mechanics of the conditioning process available. Both the degree of hydration and fines content of a sludge stream can be materially increased by exposure to shear, heat, or storage. For example, pipeline transport of sludge to central processing facilities, weekend storage of sludge prior to mechanical dewatering, and storage of sludges for long periods of time have been shown to increase the demand for conditioning chemicals prior to all types of dewatering. These factors should be considered in the design of the complete dewatering facility (1).

5.3 Inorganic Chemical Conditioning

Inorganic chemical conditioning is associated principally with vacuum and pressure filtration dewatering processes. The chemicals normally used in conditioning municipal wastewater sludges are lime and ferric chloride. Less commonly, ferrous sulfate, ferrous chloride, and aluminum sulfate have been used.

5.3.1 Ferric Chloride

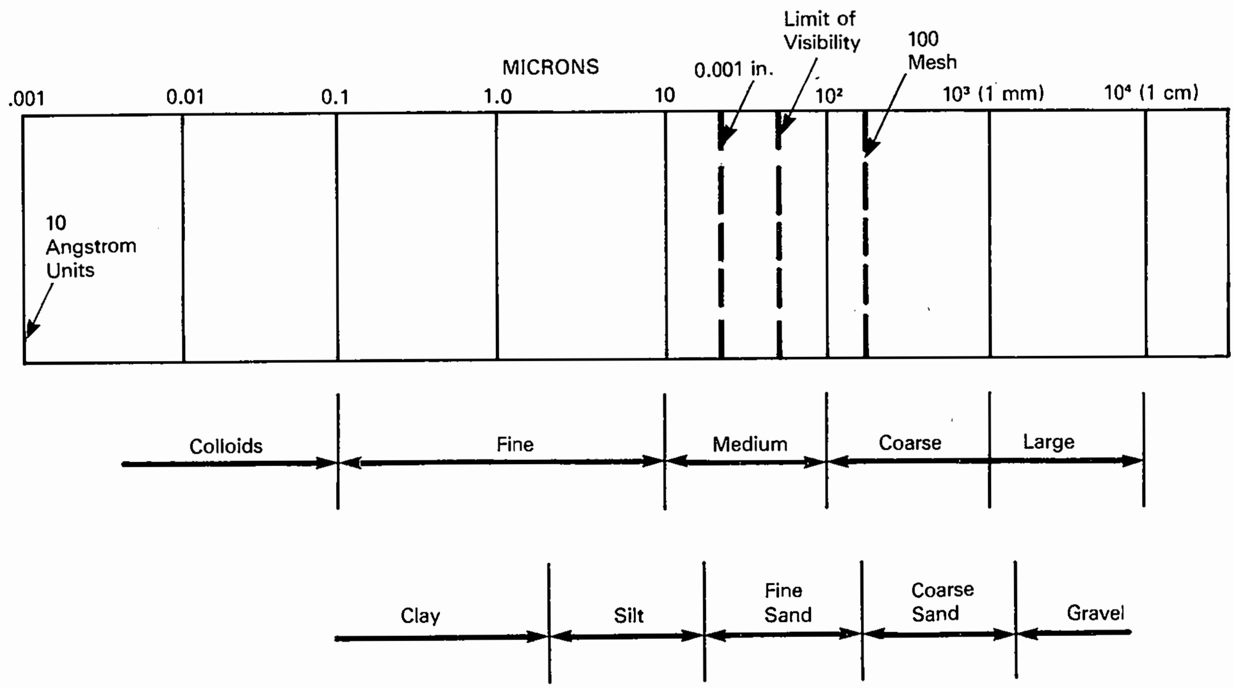
Ferric chloride is added to sludge in conjuction with lime and is added first. It hydrolyzes in water, forming positively charged soluble iron complexes that neutralize the negatively charged sludge solids, thus causing them to aggregate. Ferric chloride also reacts with the bicarbonate alkalinity in the sludge to form hydroxides that act as flocculants. The following equation shows the reaction of ferric chloride with bicarbonate alkalinity:

$$2FeCl_3 + 3Ca(HCO_3)_2 \rightarrow 2Fe(OH)_3 + 3CaCl_2 + 6CO_2 \quad (5\text{-}1)$$

Ferric choride solutions are generally used at the concentration received from the supplier (30 to 40 percent) because dilution can lead to hydrolysis reactions and the precipitation of ferric hydroxide.

An important consideration in the use of ferric chloride is its corrosive nature. Special materials must be used in its handling, with recommended materials being epoxy, rubber, ceramic, PVC, and vinyl. Contact with skin and eyes must be avoided. Rubber

Figure 5-1. Particle size distribution of common materials.



gloves, face shields, goggles, and rubber aprons should be used at all times.

Ferric chloride can be stored for long periods of time without deterioration. Usually it is stored in above-ground tanks constructed of resistant plastic or in lined steel tanks. At low temperatures, ferric chloride can crystallize, which generally means that tanks must be stored indoors or must be heated. Table 5-2 (1) shows the freezing temperature of various concentrations of ferric chloride.

Table 5-2. Crystallization Temperatures for Ferric Chloride Solutions

Solution Strength	Freezing Temperature of an Unagitated Solution	
% $FeCl_3$	°C	°F
20	-21	-5
40	-23	-10
45	-1	+30

5.3.2 Lime

Hydrated lime is usually used in conjunction with ferric iron salts. Although lime has some slight dehydration effects on colloids, it is chosen for conditioning principally because it provides pH control, odor reduction, and disinfection. $CaCO_3$, formed by the reaction of lime and bicarbonate, provides a granular structure that increases sludge porosity and reduces sludge compressibility.

Lime is available in two dry forms, quicklime (CaO) and hydrated lime ($Ca(OH)_2$). When using quicklime it is usually first slurried with water, which converts it to calcium hydroxide prior to adding it to the sludge. This process, which is called slaking, produces heat and thus special equipment is required. Quicklime normally is available in three grades: high - 88 to 96 percent CaO; medium - 75 to 88 percent CaO; and low - 50 to 75 percent CaO. These grades can affect the slaking ability of the material and should be considered when deciding which grade to purchase. In general, only quicklime that is highly reactive and quick slaking should be used for conditioning. Quicklime must be stored in a dry area, since it reacts with moisture in the air and can become unusable (2).

Hydrated lime is much easier to use since it does not require slaking, mixes easily with water (with very little heat produced), and does not require any special storage conditions. However, it is more expensive and less available than quicklime. Thus, the general rule of thumb is to obtain and slake quicklime for applications that require more than 1-2 tons per day.

5.3.3 Dosage Requirements

Iron salts, such as ferric chloride, are usually added at a dose rate of 20 to 62 kg/Mg (40 to 125 lb/ton) of dry solids in the sludge feed, whether or not lime is used. Lime dosage usually varies from 75 to 277 kg/Mg (150 to 550 lb/ton) of dry solids dewatered. Table 5-3 (3) lists typical ferric chloride and lime dosages for various sludges.

Inorganic chemical conditioning increases sludge mass. A designer should expect one pound of additional sludge for every pound of lime and ferric chloride added (1). This increases the amount of sludge for disposal and lowers the fuel value for incineration. Nevertheless, use of lime can be beneficial because of its sludge stabilization effects.

5.3.4 Design Example

A designer has calculated that the rotary drum, cloth belt vacuum filter that will be used at the plant must be capable of dewatering a maximum of 272 kg/hr (600 lb/hr) of sludge. The sludge will be a mixture of 40 percent primary and 60 percent waste activated sludge, and it will be anaerobically digested. The vacuum filter is to operate 7 hours per day, 5 days per week.

To design for a margin of safety in the chemical feed equipment, the designer has used the higher values shown in Table 5-3. Chemical feeders should be capable of adding 60 kg/Mg (120 lb/ton) of $FeCl_3$ and 210 kg/Mg (420 lb/ton) of CaO.

- Maximum daily amount of sludge to be dewatered for this example is (refer to Appendix A for examples of the calculations used to determine this value):

  272 kg sludge/hr x (7 hr/d) = 1904 kg/d (4,200 lb/d)

- Maximum amount of $FeCl_3$ required per day is:

  1,904 kg sludge/d x 60 kg $FeCl_3$/1,000 kg sludge = 114 kg/d (252 lb/d)

- The $FeCl_3$ is available as a 40 percent solution. That is it contains 1.0 kg of active ingredient per 1.77 liters of solution (4.72 lb/gal of solution).

  114 kg/d x 1.77 liters of product/1.0 kg $FeCl_3$ = 202 liters of $FeCl_3$ solution needed/day (53.4 gal/d)

- Maximum amount of CaO required per day is:

  1,904 kg sludge/d x 210 kg CaO/1,000 kg sludge = 400 kg CaO/d (882 lb/d)

- The quicklime is available at 90 percent CaO:

  400 kg CaO/d x 1 kg quicklime/0.9 kg CaO = 445 kg quicklime/d (980 lb/d)

  The amount of extra sludge produced due to chemical addition is estimated at one kg for every kg of $FeCl_3$ and quicklime added. Therefore, total maximum daily dry solids to be disposed of are:

  1,905 kg sludge + 114 kg $FeCl_3$ + 445 kg quicklime = 2,464 kg (5,432 lb) of solids

  This is the equivalent of 12,320 kg (27,160 lb) of wet solids at a minimum of 20 percent solids.

- Cost associated with this amount of chemicals in 1986 dollars:

  $FeCl_3$ = \$0.26/kg (\$0.12/lb)
  quicklime = \$0.07/kg (\$0.03/lb)

  114 kg $FeCl_3$/d x \$0.26/kg = \$29.64/d

  445 kg quicklime/d x \$0.07/kg = \$31.15/d

  1,905 kg sludge/d ÷ 1 Mg/1,000 kg = 1.9 Mg (2.1 tons)/d

  [(\$29.64 + \$31.15)/d] ÷ (1.9 tons/d) = \$31.99/dry Mg (\$29.02/ton)

5.3.5 Other Types of Inorganic Conditioners

Other types of inorganic materials have been used to condition sludge. The following is a brief description of some of these materials and their uses:

- *Coal* -- Pulverized coal has been used successfully as a conditioning agent in centrifuge and vacuum filter studies done by EPA and others (4). The recent study by Albertson and Koppers (5) showed that in a concurrent, solid bowl centrifuge, cake solids were increased from 7 to 14 percent with fine coal addition in the ratio of 0.1 to 0.3 kg coal/kg dry sludge solids (0.1 to 0.3 lb/lb). The main benefit of fine coal addition centrifuge feed seems to be the improvement in the cake solids concentration. Because of the increased moisture removal provided by the fine coal feed, fuel costs for sludge combustion can be reduced as much as 60 to 90 percent. Some concerns, including those of safety, however, have arisen concerning materials handling, dust generation, and incinerator temperature control with the addition of coal.

- *Cement Kiln Dust* -- Cement kiln dust has been used to successfully condition sludge prior to dewatering on vacuum filters and also for before and after stabilization. Kiln dust is a byproduct of the cement and lime industries and is high in calcium and potassium. About twice the amount of kiln dust is required to achieve the same pH as from lime. However, the cost is reported to be about 30 percent that of lime. Some material handling problems have been reported, but the

Table 5-3. Typical Conditioning Dosages of Ferric Chloride and Lime for Municipal Wastewater Sludges*

Type of Sludge	Vacuum Filter		Pressure Filter	
	$FeCl_3$	CaO	$FeCl_3$	CaO
Raw:				
Primary	40-80	160-200	80-120	20-280
WAS	120-200	0-320	140-200	400-500
Primary + TF	40-80	180-240		
Primary + WAS	50-120	180-320		
Primary + WAS (septic)	50-80	240-300		
Elutriated Aerobically Digested:				
Primary	50-80	0-100		
Primary + WAS	60-120	0-150		
Anerobically Digested:				
Primary	60-100	200-260		
Primary + WAS	60-120	300-420		
Primary + TF	80-120	250-350		
Thermally Conditioned:	None	None	None	None

* All values shown are for pounds of either $FeCl_3$ or CaO per ton of dry solids pumped to the dewatering unit.
lb/ton x 0.5 = kg/Mg

advantages appear to warrant further investigation (6).

- *Ash* -- Flyash, power plant ash, and sludge incinerator ash can be used as sludge conditioning agents to increase a sludge's dewatering rate, improve cake release, increase cake solids, and in some cases reduce the dosage of other types of conditioning agents. As early as 1927 a process was patented for taking sludge incinerator ash from an incineration unit back to a vacuum filter to assist dewatering. Ash has been used both as a precoat and as a body feed in one manufacturer's dewatering system with high pressure filtration. However, usually ferric chloride and lime must be used with flyash. The City of Indianapolis, Indiana has successfully used ash as a conditioner on their rotary belt vacuum filters to minimize conditioner requirements and enhance cake release from the media.

5.4 Organic Polymers

During the past decade, important advances have been made in the manufacture of polymers for use in wastewater sludge treatment. Polymers are now widely used in sludge conditioning and a large variety are now available. It is important to understand that these materials differ greatly in chemical composition, functional effectiveness, and cost effectiveness.

Polymers were originally used to condition primary sludges and easy-to-dewater mixtures of primary and secondary sludges for dewatering by rotary vacuum filters or solid bowl decanter centrifuges. Improvements in the effectiveness of polymers has led to their increasing use with all types of dewatering processes. Reasons for selecting polymers over inorganic chemical conditioners are:

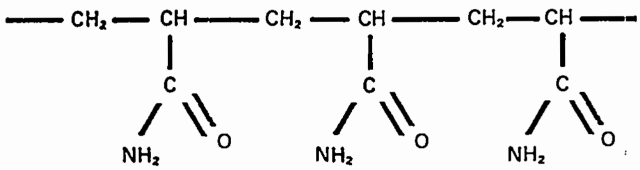
- Little additional sludge mass is produced. Inorganic chemical conditioners typically increase sludge mass by 15 to 30 percent.
- If dewatered sludge is to be used as a fuel for incineration, polymers do not lower the fuel value.
- They allow for cleaner material-handling operations.
- They reduce operation and maintenance problems.

Selection of the correct polymer requires that the designer work with the polymer suppliers, equipment suppliers, and plant operating personnel. Evaluations should be made on site and, if possible, with the sludges to be conditioned. Since new types and grades of polymers are continually being introduced, the evaluation of polymers must be an ongoing process.

5.4.1 Composition and Physical Form

Polymers are long chain, water soluble specialty chemicals. They can be either completely synthesized from individual monomers, or they can be made by the chemical addition of functional monomers or groups to naturally occurring polymers. A monomer is the subunit or repeating unit from which polymers are made through various types of polymerization reactions. The backbone monomer most widely used in synthetic organic polymers is acrylamide. Polyacrylamide, created when the monomers combine to form a long, thread-like molecule with a molecular weight in the millions, is shown in Figure 5-2. In the form shown polyacrylamide is essentially non-ionic. That is to say it carries no net electrical charge in aqueous solutions. However, under certain conditions and with some solids, the polyacrylamide can be sufficiently surface active to perform as a flocculant. In addition, it is often used as a flocculation aid in conjunction with lower molecular weight primary coagulants which are charged molecules.

Figure 5-2. Polyacrylamide molecule-backbone of the synthetic organic polymer.



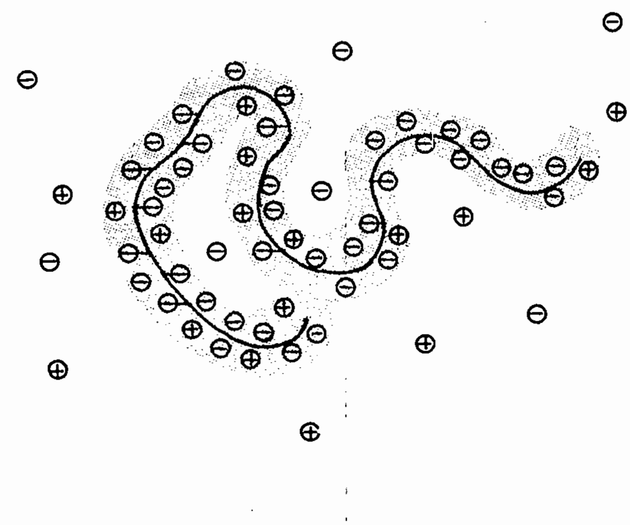
Anionic-type polyacrylamide flocculants carry a negative electrical charge in aqueous solutions and are made by either hydrolyzing the amide group (NH_2) or combining the acrylamide monomer with an anionic monomer. Cationic polyacrylamides carry a positive charge in aqueous solutions and can be prepared by chemical modification of essentially non-ionic polyacrylamide or by combining a cationic monomer with acrylamide. When cationic monomers are copolymerized with acrylamide in varying proportions, a family of cationic polymers with different degrees of charge and molecular weights are produced. These polymers are the most widely used polymers for sludge conditioning, since most sludge solids carry a negative charge. The characteristics of the sludge to be processed and the type of dewatering device used will determine which of the cationic polymers will work best and still be cost-effective. For example, an increasing degree of charge is required when sludge particles become finer, when hydration increases, and when relative surface charge increases.

Polymers are available as dry powders or liquids. The liquids come as water soluble solutions or as water-in-oil emulsions. The shelf life of dry powders is usually one or more years, whereas most of the liquids have shelf lives of about 6-12 months and must be protected from wide ambient temperature variation in storage. Polymers can also be purchased with various molecular weights and charge densities which can greatly affect the conditioning characteristics of the polymer and its reaction with the sludge.

5.4.2 Structure in Solution

Organic polymers dissolve in water to form solutions of varying viscosity. The resulting viscosity depends on their molecular weight, degree of ionic charge and salt content of the dilution water. At infinite dilution, the molecule tends to assume the form of an extended rod because of the repulsive effect of the adjacent, charged sites along the length of the polymer chain. At normal concentrations the long thread-like charged anionic polymer assumes the shape of a random coil, as shown in Figure 5-3.

Figure 5-3. Typical configuration of an anionic polymer in solution.



This simplified drawing, however, neither shows the tremendous length of the polymeric chain, nor the very large number of active polymer chains that are available in a polymer solution. It has been estimated that a dosage of 0.2 mg/l of polymer having a molecular weight of 100,000 would provide 120 trillion active chains per liter of water treated.

Dewatering is inhibited by the physical and chemical characteristics of the sludge particles. Polymers in solution act by adhering to the sludge particle, causing the following phenomena to occur (see Figure 5-4):

- Desorption of bound surface water
- Charge neutralization
- Agglomeration of small particulates by bridging between particles.

5.4.3 Dry Polymers

Representative dry polymers are described in Table 5-4. This table does not list the myriad of available types and proprietary chemical differences capable of yielding performance advantages in different sludge systems, but it does show some of the gross distinctions among the major types.

Dry polymers are available in powdered, granular, bead, or flake form. The form is usually determined by the manufacturing process. Dry polymers have

Figure 5-4. Schematic representation of the bridging model for the destabilization of colloids by polymers.

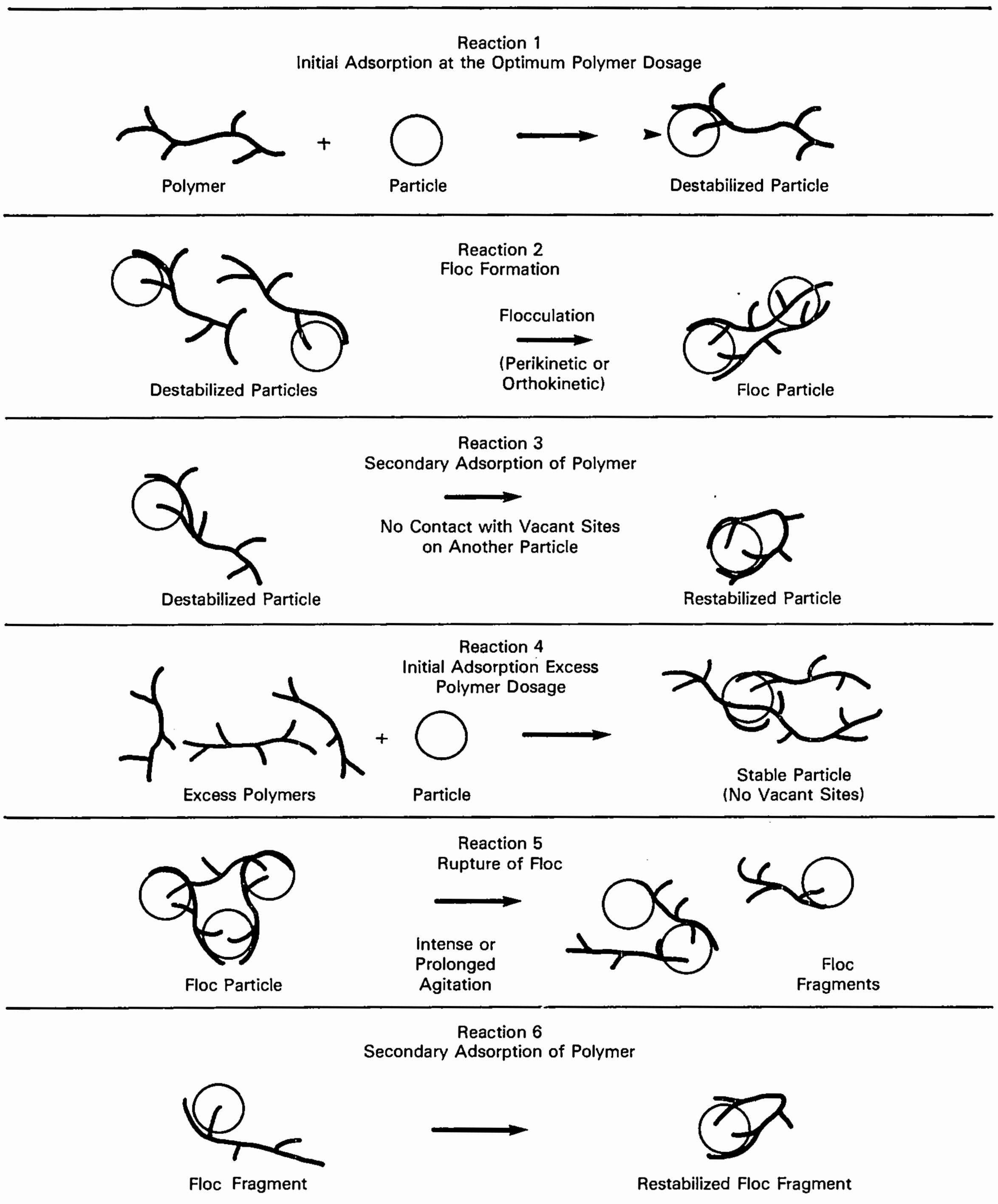


Table 5-4. Representative Dry Powder Cationic Polymers (Polyacrylamide Copolymers)

Relative Cationic Density[1]	Molecular Weight[2]	Approximate Dosage
		lb/ton dry solids
Low	Very high	0.5-10.0
Medium	High	2.0-10.0
High	Medium high	2.0-10.0

[1] Low: < 10 mole %
Medium: 10-25 mole %
High: >25 mole %
[2] Very High: 4,000,000-8,000,000
High: 1,000,000-4,000,000
Medium high: 500,000-1,000,000
lb/ton x 0.5 = kg/Mg

very high activities (the amount of polymeric chemical contained in the product). The active solids concentration is usually as high as 90 to 95 percent. Dry polymer should be stored in a cool, dry area and should not be exposed to moisture, since it will tend to cake the polymer and make it unusable.

Dissolving dry polymer requires care. A typical dry polymer make-up system, shown in Figure 5-5 (7), contains several important components. The system should include an eductor or other polymer wetting device that allows for the proper pre-wetting of the polymer particles before they enter the mix tank. After the pre-wetted polymer enters the tank, it should be mixed slowly until it is completely dissolved. Mixing should then continue for at least 60 minutes to insure that all of the polymer is completely dissolved. Undissolved polymer can cause many problems, including clogging of pumps and piping and fouling of the filter belts and cloths. Mixing also allows time for the polymer to age. During the aging process, the molecule uncoils and takes on a form that enables it to cause flocculation of the sludge. If the polymer solution is not allowed to age, the polymer will not perform as expected.

5.4.4 Liquid Polymers

The various liquid cationic polymers, which are either concentrated water solutions or emulsions suspended in hydrocarbon oils, are described in Table 5-5. Liquid polymers are available in various activities (percent active solids). The concentration of the polymeric material that the manufacturer can dissolve in water is usually controlled by the viscosity of the final solution.

Water solutions of polymer are usually purchased in either 208-liter (55-gal) drums, in liquid bins of about 1,040 liters (275 gal), or in bulk quantities of about 19,000-23,000 liters (5,000 to 6,000 gal). In colder climates, storage areas for drums, liquid bins and bulk liquid should be located indoors and heated. If the bulk storage tank must be located outdoors, the tank should be heated so that the solution's viscosity will be low enough to allow pumping. Polymers form true solutions. Thus, mixing of the concentrated polymer is not required.

The bulk storage tank should be fitted with a sight-glass and low and high level sensors. The low level sensor can be wired into a control panel in the operations area. This sensor can be set to sound an alarm when there is a certain amount of polymer remaining in the tank, enabling plant personnel to re-order polymer before running out of material. For example, if it takes 10 days to receive a shipment of polymer from the day of order, the low level sensor should be set to ring when a 10-day supply is reached. The high level sensor is important during the delivery operation to insure that no polymer is spilled. Polymer is extremely difficult to clean up and precautions must be taken to prevent spills.

To safely convey the polymer from drums into the mix tank, the operator should use either a polymer transfer pump or a drum lifting device to empty the drum. Polymer transfer pumps are also required to pump the material from the storage tank to the mix tank. They should be of the progressive cavity type so that the polymer molecule is not subjected to high shear forces.

It is also advisable to include a timer on the pump control panel. The operator can then set the timer to a specified interval, and thus the pump will always transfer the exact amount of polymer to the mix tank. The pump should be calibrated on a regular basis (monthly, for example) to insure that the same quantity of polymer is being pumped in that time interval. The designer should insure that the timer setting can be easily changed.

The preparation system, as shown in Figure 5-6 (7), for this type of polymer should include a mixing tank and a storage tank for the diluted polymer. Typically, the operator will prepare a 0.1 percent solution of polymer. The concentrated polymer and water should be mixed for about 30 minutes to insure a homogeneous solution. Once the polymer has been diluted, it is usually stable for about 24 hours. Therefore, only enough polymer should be made up to use in that period.

Emulsions are dispersions of polymer particles in a hydrocarbon oil. Surface active agents are used to prevent separation of the polymer-oil phase from the water phase. Activities as high as 25 to 50 percent are common with emulsions. Emulsions are available in 208-liter (55-gal) drums, in 1,040-liter (275-gal) liquid bins, or in bulk quantities. Storage requirements are the same as for water solutions of polymer, except care must be taken that no water comes in contact with the emulsion until it is ready for

Figure 5-5. An example of a dry polymer make-up system.

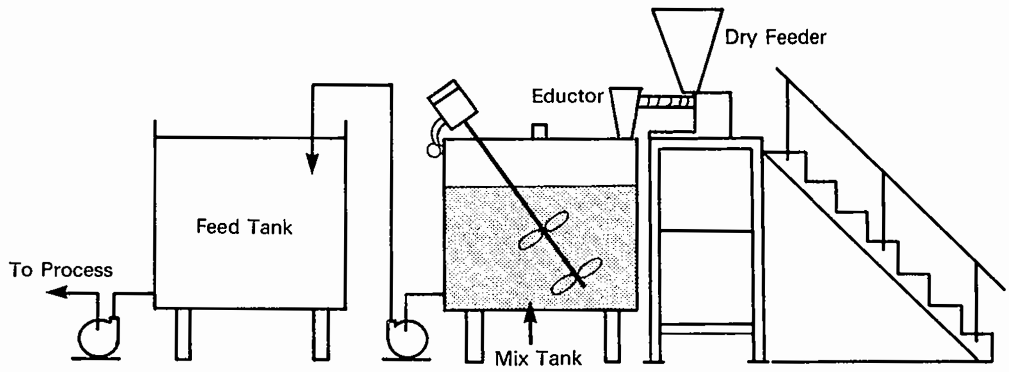
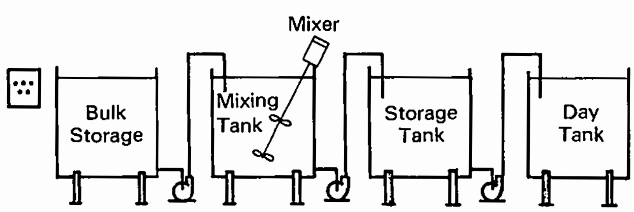


Table 5-5. Representative Cationic Polymers

Type	Relative Cationic Density[a]	Molecular Weight[b]	Percent Solids
Solution type:			
Mannich product[c]	High	High to very high	4-8
Tertiary polyamine	High	Low	20-50
Quaternary polyamine	High	Very low to medium	20-50
Quaternary poly DADM	High	Low to medium	20-40
Emulsion type:			
Polyacrylamide copolymer	Low Medium High	High to very high	25-60

[a] Tertiary amines charge affected by solution pH; lose cationically in alkaline environments.
[b] Very low: < 100,000
Low: 100,000-200,000
Medium: 200,000-1,000,000
High: 1,000,000-4,000,000
Very high: 4,000,000-8,000,000
[c] Product of polymerization reaction of condensation type. Specifically, the condensation reaction of a primary or secondary amine with formaldehyde and a ketone to form a beta amino ketone.

Figure 5-6. An example of a liquid polymer make-up system.



mixing and that the temperature of the storage area is fairly constant. Premature exposure to water will cause the polymer to coagulate. Mixing and aging of the emulsion polymers also requires care. An emulsion polymer make-up system is shown in Figure 5-7 (7). Compact and portable polymer feed automation equipment is available for in-line use that requires no batch mixing or aging tanks.

Initially, the emulsion must be broken. Usually, a disperser uses high pressure water to contact and break the emulsion. The polymer should be aged for at least 30 minutes after it has been diluted before use. Follow all manufacturer's recommendations for mixing and aging of the emulsion to insure optimum performance.

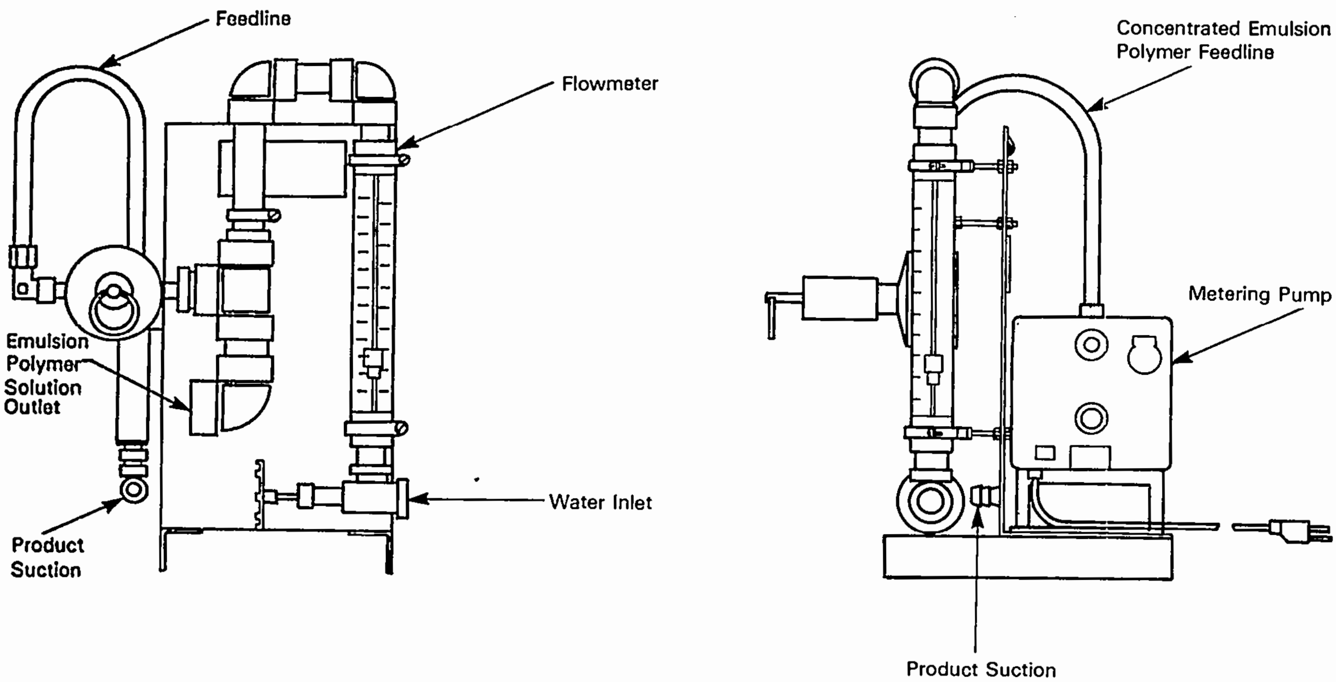
5.4.5 Polymer Feed

A typical polymer feed system should include a day tank, polymer feed pumps, dilution water system, and alternate feed points to the dewatering units. It can also include an in-line static mixer. The day tank should be sized to hold a 1-day supply of diluted polymer or less, be made of fiberglass (which provides the broadest corrosion resistance) and should be equipped with a slow-speed mixer and a sight-glass or level gauge.

The feed pumps should be of the progressive cavity type to insure that the minimum amount of shear forces are exerted on the polymer. (Diaphragm pumps are also used sometimes.) These pumps should be calibrated weekly to insure accurate dosages of polymer. By using variable speed pumps, the operating personnel can adjust the polymer dosage to compensate for changes in sludge characteristics. It is important to note that over-conditioning the sludge is just as bad as under-conditioning and will produce a sludge that is very difficult to dewater. Polymer dosages should be re-evaluated periodically (see Section 5.5.3).

In-line dilution water is a very important part of polymer use. This water further dilutes the polymer and makes it disperse more readily in the sludge, thus conditioning the sludge more effectively. Polymer manufacturers supply dilution information on the type of water systems required. Typically 4 to 15 l/min (1 to 4 gpm) of dilution water is needed and depends on the polymer feed rate.

Figure 5-7. An example of an emulsion polymer make-up system.



The location of polymer feed points can greatly affect the performance of the polymer and therefore, the dewatering unit. For centrifuges, the polymer feed point is usually inside the dewatering unit itself (see Section 7.3). However, for a belt filter press, at least two or three optional locations should be specified, one adjacent to the dewatering unit, one about 1 to 1.5 m (3 to 5 ft) upstream and tied into the sludge feed line, and one about 6 to 9 m (20 to 30 ft) upstream. Usually connecting the feed points into the sludge piping upstream of the unit works best. In this case, the sludge has more time to mix with the polymer and therefore forms a better floc. However, overmixing should be avoided since excess shear can degrade fragile floc.

5.4.6 Typical Polymer Dosages

5.4.6.1 Belt Filter Presses

Compared to other mechanical dewatering processes, belt filter presses appear to have the greatest need for optimizing the polymer dosage as a function of the incoming sludge's characteristics (8). Under-conditioning results in inadequate drainage of the free water in the gravity dewatering zone, which can cause sludge overflow in the gravity zone or extrusion of the sludge in the pressure section. Under-conditioned biological solids can also blind or clog the filter belt. Over-conditioning can cause belt blinding and problems with cake release from the belt by making the sludge sticky. Futhermore, overflocculated sludge may drain so rapidly that the solids are not distributed evenly across the media. Uneven distribution can cause tracking problems with the belt and, moreover, can produce poor quality sludge cake. Table 5-6 (3) lists typical levels of dry polymer addition to condition sludge for dewatering on a belt filter press.

Table 5-6. Typical Dosages of Dry Polymer for Belt Filter Presses

Type of Sludge	Pounds of Dry Polymer per Ton of Dry Solids	Typical Values	Cost per Dry Ton*
Raw:			
Primary	2-9	5	4.30-19.35
Primary + TF	3-15	10	6.45-32.25
Primary + WAS	2-20	7	4.30-43.00
WAS	2-20	10	4.30-43.00
Aerobically Digested:			
Primary + WAS	4-15	10	8.60-32.25
Anerobically Digested:			
Primary	2-10	3	4.30-21.50
Primary + WAS	3-15	6	6.45-32.25

* $2.15/lb (1986 cost for dry polymer)
lb/ton x 0.5 = kg/Mg

5.4.6.2 Solid Bowl Centrifuges

Solid bowl centrifuges usually require polymer to obtain good performance on municipal wastewater sludges. Table 5-7 (3) lists typical levels of dry polymer dosages to various sludges for conditioning prior to dewatering in a centrifuge.

Table 5-7. Typical Dosages of Dry Polymer for Conditioning Various Types of Sludges for Dewatering in Solid Bowl Centrifuges

Type of Sludge	Pounds of Dry Polymer per Ton of Dry Solids	Typical Values	Cost per Dry Ton*
Raw:			
Primary	2-7	4	4.30-15.50
Primary + WAS	4-15	8	8.60-32.25
Anerobically Digested:			
Primary	6-10	6	12.90-21.50
Primary + WAS	7-15	8	15.50-32.25
Thermally Conditioned:			
Primary + WAS	3-5	3	6.45-10.75
Primary + TF	7-15	8	15.50-32.25

* $2.15/lb (1986 cost for dry polymer)
lb/ton x 0.5 = kg/Mg

5.4.6.3 Vacuum Filters

Many of the vacuum filter installations in the United States have now converted from ferric chloride and lime as the conditioner to polymer. With polymer there are many advantages, such as lower costs and fewer material handling problems. Further, the mass of solids to be disposed of will not increase as occurs with inorganic conditioners, and the volatile content of the sludge cake will be higher. Table 5-8 (3) shows amounts of dry polymer to condition different types of sludge for vacuum filtration.

Table 5-8. Typical Dosages of Dry Polymer for Conditioning Various Types of Sludges on Vacuum Filters

Type of Sludge	Pounds of Dry Polymer per Ton of Dry Solids	Typical Values	Cost per Dry Ton*
Raw:			
Primary	0.5-1	1	1.08-2.15
Primary + TF	2.5-5	4	5.38-10.75
Primary + WAS	4-10	6	8.60-21.50
WAS	8-15	12	17.20-32.25
Anerobically Digested:			
Primary	1.5-4	1.5	3.23-8.60
Primary + WAS	5-12	7	10.76-25.80

* $2.15/lb (1986 cost for dry polymer)
1 lb/ton = 0.5 kg/metric ton

5.4.6.4 Drying Beds

Sludge added to drying beds can be conditioned with polymer. Indications are that adding 0.25 to 1.0 kg of dry polymer/Mg of dry solids (0.5 to 2.0 lb/ton) can increase dewatering rates. Chapter 8 provides case studies with examples of quantities of polymer to be added to drying beds.

5.4.6.5 Pressure Filters

Engineers and operators at pressure filter installations are experimenting with polymers to replace inorganic conditoners. Some of the newer polymers on the market appear to give good performance on the filters. Advantages of using polymers would be lower costs, reduced material handling, and no increase in sludge mass for final disposal.

5.5 Design of a New Installation

The first design step is to determine how much and what type of sludge will be produced at the treatment plant. Calculations such as those shown in Section 7.2.6 can be used to predict, in general, the quantity of sludge that will be produced. Another method of estimating sludge quantity is to use the following averages:

- 3 liters of 4-percent solid, primary sludge are produced per m^3 of wastewater treated (2,980 gal/10^6 gal);
- 18 liters of 1-percent solid, waste activated sludge are produced per m^3 of wastewater treated (18,025 gal/10^6 gal).

These numbers were caculated by averaging values from four different authors (9). The designer must take care when using either the generalized equations or these averages, since they do not take into account unusual conditions in a particular community that could impact on the sludge quantity. If at all possible, the designer should try to determine actual production rates by performing pilot plant studies of the waste.

The type of sludge produced is determined by the wastewater's characteristics and the wastewater treatment process, i.e., activated sludge, trickling filter, anaerobic digestion, etc. Each type of sludge has different conditioning requirements.

Once these parameters are established, the engineer can design the required equipment.

5.5.1 Design Example

For this example, assume a 18,927-m^3/d (5-mgd) activated sludge treatment plant with primary clarifiers. After conditioning with polymer, the sludge will be fed to belt filter presses.

Using the averages shown above, the plant will be producing 56,397 liters (14,900 gal) of 4-percent primary sludge and 341,123 liters (90,125 gal) of 1-percent secondary sludge. If the mixture of sludges were thickened to 5 percent solids, then 113,346 liters (29,946 gal) of mixed sludge or 5.6 dry Mg (6.2 tons) would have to be dewatered per day. Assuming that dewatering is to take place on a 5-day week, 7.9 dry Mg (8.7 tons) must be processed per day.

Referring to Table 5-6, 0.9 to 9 kg (2 to 20 lb) of polymer are required per dry ton of solids. Assuming a dry polyacrylamide polymer and using the highest value of 10 kg/Mg (20 lb/ton), the maximum daily polymer use will be about 78 kg/d (174 lb/d). Based on this, the designer must then calculate the volume of the mixing tank, storage tank, and day-tank as well as the capacity of the various transfer pumps. He must also calculate the amount of storage space for the dry polymer (number of 55-gallon drums or volume of the bulk storage tank, depending on what form of polymer is to be used). The bulk polymer tank should be sized for not more than a 6- to 8-week supply. This will insure that the polymer is always fresh. As an alternative, it may be convenient to specify automated polymer blending and feeding equipment.

5.5.2 Additives

Potassium permanganate, which is often used for odor control, has been shown to reduce polymer doses on mechanical dewatering units (R. McRoberts, Carus Chemical, LaSalle, IL, personal communication, 1982). Treatment plant operating personnel, who have optimized the polymer use in conjunction with permanganate use, report a 5 to 15 percent reduction of polymer.

5.5.3 Selection of a Conditioning Chemical

Many factors go into the selection of the appropriate conditioning chemical to be used at a particular plant. These factors include such considerations as performance, material handling, storage requirements, type of dewatering units, final disposal method and economics. For example, a plant whose final disposal method is incineration wants the driest cake possible with the least mass and highest volatile content. Therefore, polymer conditioning is usually the better choice when compared to inorganic chemicals. Polymer conditioning also proves to be the better choice if either storage space is at a minimum or if material handling could be a problem.

Today, most municipal wastwater treatment plants are selecting polymer conditioning over inorganic chemical conditioning. Manufacturers' representatives can be of great assistance in evaluating conditioning agents. However, there are several tests, which the designer can perform quickly and inexpensively, that will provide a great deal of information about the conditioner's performance. Such tests can also estimate the quantities of agents that will be required. If sludge is not available from the plant, the designer could use either pilot-plant sludge or sludge from a similar plant. However, once the plant is on-line, the conditioner must then be re-evaluated.

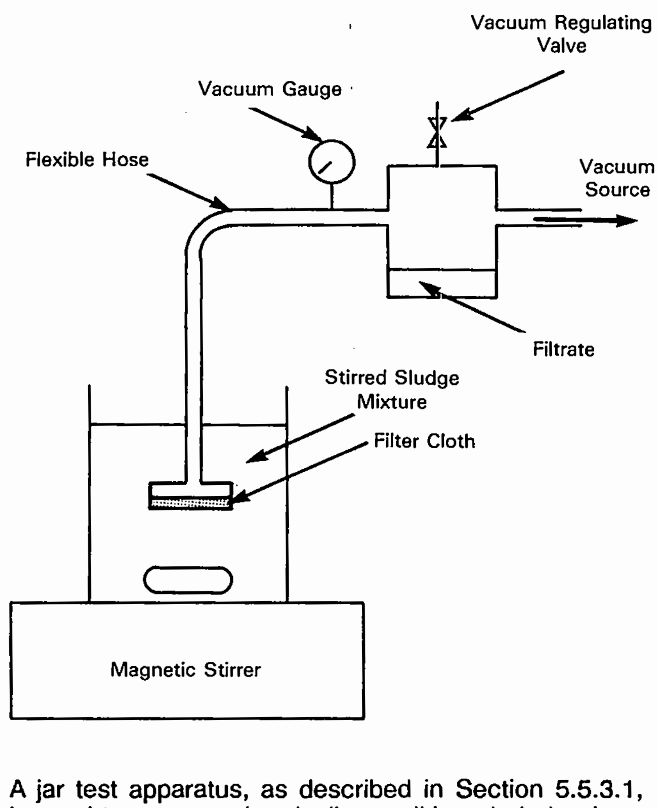
5.5.3.1 Jar Test (10)

The Jar Test is used to screen conditioning agents especially when the designer faces a wide variety of potentially effective products. This test is performed by taking four to six large beakers of about 1-liter capacity and filling them with about 600 ml of the sludge. Solutions of different types of conditioning chemicals are prepared in accordance with the manufacturer's instructions. Conditioning chemicals could include ferric chloride, lime, and up to about four polyelectrolytes. Each of the conditioning chemicals (ferric chloride, the four polyelectrolytes, and ferric chloride and lime in tandem) can then be added to a different sludge sample at the manufacturer's suggested dosage levels or at levels noted previously in this chapter. The beakers are then placed on a gang stirrer with the filter pan in position beneath the oversized paddle. The paddle should just clear the bottom of the pan. The stirrer should be set to 75 rpm. The diluted chemical is then poured into the filter pan and mixed for 30 seconds. The operator then stops the gang stirrer, removes the paddle, and observes the floc formation and settling.

5.5.3.2 Filter Leaf Testing

The Filter Leaf Test (1,9) is usually used for evaluating dewaterability, primarily by a vacuum filter. Further, in some cases, this test has been used to size a vacuum filter. The test is performed by assembling a filter leaf apparatus as show in Figure 5-8. The filter cloth should be the fabric intended for use or monofilament filter cloth.

Figure 5-8. Example of a filter leaf apparatus.



A jar test apparatus, as described in Section 5.5.3.1, is used to prepare chemically conditioned sludge in at least two-liter batches for each filtration cycle. The conditioning chemicals are placed in the jar test apparatus, allowing for 2 to 4 minutes of mixing and flocculation time. The mixing should be slow (about 10 rpm). Flash mixing will adversely affect test results.

Two liters of the chemically conditioned sludge are transferred to a beaker, its temperature is measured and the filter leaf is submerged in it about 5 cm (2 in) below the surface. Then 51 cm (20 in) Hg of vacuum is applied to the filter leaf and timing begins. After 45 seconds of form time, the leaf is withdrawn and dried for 90 seconds.

The cake thickness is measured and the cake is scraped into a previously weighed dish. The dish and cake are weighed and transferred to a drying oven. After air drying, the cake should be desiccated, weighed, volatilized, desiccated again, and weighed again.

The following determinations should be made for each run:

- Volume of filtrate, ml
- Temperature of filtrate, °C
- Wet weight of filter cake, g
- Dry weight of filter cake, g
- Dry weight of ash, g
- Total solids concentration of cake, Ts, percent of wet weight
- Volatile solids concentration of cake solids, Vs, percent of total solids

This test simulates a 3-minute cycle time, divided into 45 seconds of form time, 90 seconds of drying time, and 45 seconds of discharge time. It may be desirable to use a longer cycle time. Typically, cake thickness with cycle time and cake total solids content will increase. Cake solids content can be further increased by decreasing the ratio of form time to drying time. Subsequently, experiments might be conducted to determine yield as a function of percent solids concentration and/or a function of cycle time. Also for each run the filter yield, in pounds of cake solids per square foot of filter per hour, can be calculated with Equation 5-2:

$$\text{Filter Yield} = \text{Dry Weight of Cake (g)} \times \text{Cycles/hr} \div [453.6 \times \text{Filter Area } (\text{ft}^2)] \qquad (5\text{-}2)$$

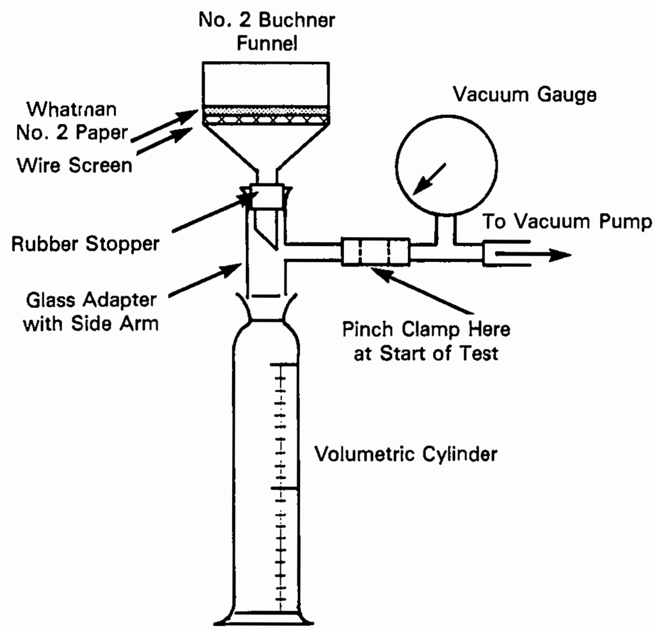
One of the advantages of the filter leaf test is that it simulates actual behavior on a vacuum filter. Ease of cake release from the filter cloth can be estimated, and the percent moisture of the final cake can be determined.

NOTE: The actual sizing of a vacuum filter would require a series of leaf tests, employing one or more controls in which no characteristics are added. A range of chemical dosages and combinations should then be studied.

5.5.3.3 Specific Resistances or Buchner Funnel Test (11)

The Specific Resistance Test is another method of predicting conditioning agent performance. A detailed theoretical description is contained in section 7.4.4 of this manual. The Buchner Funnel test equipment consists of a graduated cylinder, Buchner Funnel, and a vacuum pump as shown in Figure 5-9 (12).

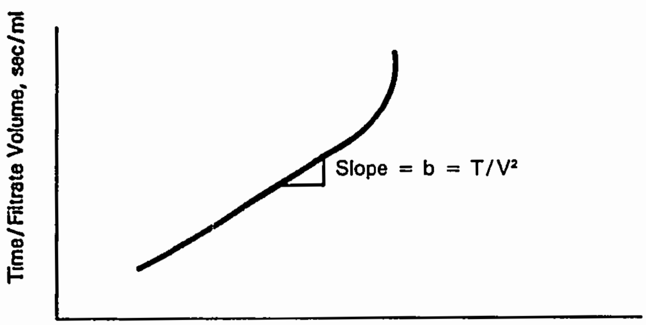
Figure 5-9. Buchner funnel apparatus.



A series of conditioned sludge samples are prepared in large beakers as previously discussed in Sections 5.5.3.1 and 5.5.3.2. First about 200 ml of the thickened sludge is placed into the beakers. The sludge tested should be representative of the sludge to be used on the dewatering units. This sludge can be from a pilot-plant or a similar full-scale treatment plant.

A Buchner funnel is mounted on top of a graduated cylinder as shown in Figure 5-9, and the funnel is fitted with a piece of filter paper. For each test, a portion of the conditioned sludge (50-200 ml) is poured into the funnel. After 2 minutes of gravity drainage, the vacuum pump is turned on (15 in Hg). At about 15-second intervals, the filtrate volume is measured and recorded until the vacuum breaks or additional water can not be removed. The sludge cake is then removed from the filter and placed in a weighed dish. The wet weight of the cake is measured and then after drying at 180°C, the dry weight is measured. Total suspended solids is determined on the filtrate sample. In addition, the temperature of the filtrate is also measured. A plot is made of time/filtrate volume versus filtrate volume, as shown in Figure 5-10. The slope of the straight line

Figure 5-10. Time/filtrate vs. filtrate volume plot.



portion of the graph is "b" and is used to calculate the specific resistance (r) from Equation 5-3:

$$r = (2\,PA^2\,b)/\mu w \qquad (5\text{-}3)$$

where,

- r = specific resistance, m/kg
- P = pressure of filtration, N/m^2
- A = area of filter, m^2
- b = slope of time/volume vs. volume curve, sec/cm^6
- μ = viscosity of filtrate, N (sec)/m^2
- w = weight of dry solids/volume of filtrate, kg/m^3

Specific resistance must be reported in the units of m/kg.

Figure 5-11 shows a plot of specific resistance versus conditioning chemical dose. This plot was constructed with specific resistance data from a sludge conditioned with different levels of the same chemical, in this case a polyelectrolyte. From a plot such as this, the designer can determine optimum polymer dose. The optimum conditioner chemical dosage is that which produces the lowest specific resistance.

A modification of the Buchner Funnel test can be used to duplicate the gravity drainage results which can be achieved on a belt filter press (BFP). This test uses the apparatus shown in Figure 5-9, exclusive of the vacuum pump. A piece of the belt material which will be used in the BFP is placed in the Buchner Funnel. A sample of conditioned sludge is placed into the Funnel and the volume of water released is measured at regular intervals. Both time and volume are recorded. The polymer and/or dose of polymer which gives the greatest volume of free water in the shortest time should give the best results on the BFP.

5.5.3.4 Capillary Suction Time

The Capillary Suction Time (CST) (13) is a simple and quick test that measures the time required for the liquid portion of the sludge to travel 1 centimeter or any other fixed distance. The apparatus (Figure 5-12) consists of a timing device, an upper plate containing probes that activate and deactivate the timing device, and a lower plate that holds the filter paper and a metal sample container.

Figure 5-11. Plot of specific resistance vs. conditioning chemical dosage.

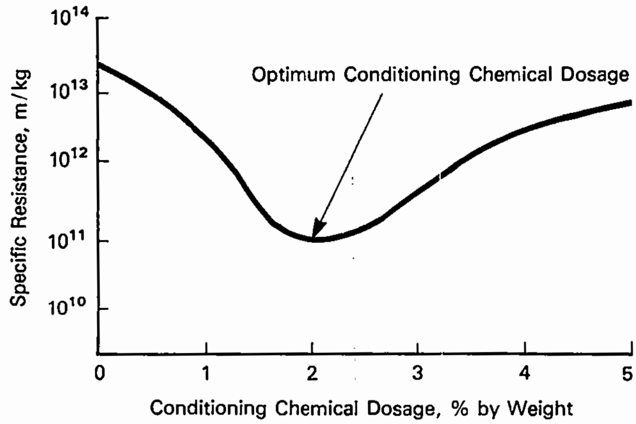
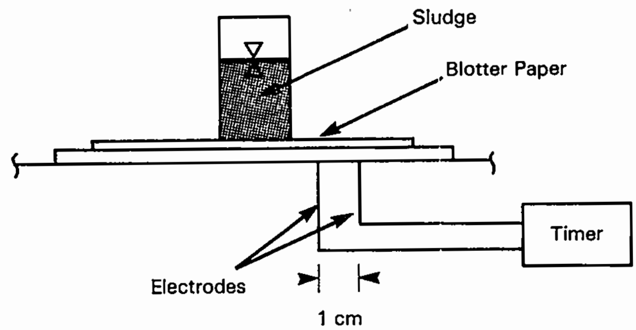


Figure 5-12. Capillary suction time apparatus.



A sample of conditioned sludge is placed in the sample container. As water migrates through the paper and reaches the first probe, it activates the timer. When the water reaches the second probe, the timer deactivates. The time interval between timer activation and deactivation is the capillary suction time and is a measure of the dewaterability of the conditioned sludge. Capillary suction time is plotted versus chemical dosage. The dosage that gives the fastest time is the optimum. Conditioner types and concentrations should be varied until the optimum chemical and dosage is found for a particular dewatering system.

5.5.4 Calculations Associated with Polymer Use

Equations 5-4 through 5-7 can help the designer to evaluate and control a polymer system.

Calculation of Dilute Polymer Concentration from A Water Solution of Polymer

$$Cn = Wp/(Vw + Vp) \qquad (5\text{-}4)$$

where,

- Cn = diluted polymer concentration, grams/liter (lb/gal)
- Wp = weight of polymer added to mixing tank per batch, g (lb)
- Vw = volume of water added to mixing tank per batch, liter (gal)
- Vp = volume of polymer added, liter (gal)

or

$$Cn = (Vp \times Dp) \div (Vw + Vp) \qquad (5\text{-}5)$$

where,

- Dp = density of concentrated polymer in grams/liter (lb/gal)

Calculation of Polymer Use and Cost per Dry Ton of Solids Dewatered

Use: $$Pt = Pu/Ws \qquad (5\text{-}6)$$

where:

- Pt = kg (lb) polymer used per dry Mg (dry ton) of dewatered sludge
- Pu = weight of polymer used per day, kg/d (lb/d)
- Ws = dry weight of sludge dewatered per day, Mg/d (tons/d)

Cost: $$Pc = Cp/Pt \qquad (5\text{-}7)$$

where,

- Pc = polymer cost per ton of sludge, $/Mg ($/dry ton)
- Cp = cost of polymer, $/kg ($/lb)

5.6 Thermal Conditioning (14)

The thermal conditioning process enhances the dewatering characteristics of sludge through the simultaneous application of heat and presssure. It is a continuous flow process in which sludge is heated to temperatures of 177°C to 204°C (350°F to 400°F) in a reactor under pressures of 1,720 to 2,750 kPa (250 to 400 psig) for 15 to 40 minutes. There are two basic modifications of the thermal conditioning process employed in wastewater treatment. In one modification, Low Pressure Oxidation (LPO), air is added to the process. The other modification, Heat Treatment (HT), does not include the addition of air to the process. Both thermal conditioning processes produce biologically stable sludge with excellent dewatering characteristics.

Wastewater sludge contains water and cellular and inert solids that form a gel-like structure. The water portion consists of bound water, which surrounds each solids particle, and water of hydration, which is inside the cellular solids. Thermal conditioning improves sludge dewaterability by subjecting the sludge to elevated temperature and pressure in a confined reactor vessel: thus coagulating solids, breaking down the gel-like structure of the sludge, and allowing the bound water to separate from the solids particles. In addition, hydrolysis of protein material in the sludge occurs. Cells break down and water is released, resulting in coalescence of solids particles. In its conditioned state, the sludge is readily dewatered on most dewatering devices to 30 to 50 percent solids, in most cases without addition of chemicals.

A portion of the volatile suspended solids (VSS) in sludge is solubilized as a result of the breakdown of the sludge structure. The solubilizaton of VSS increases its biodegradability. Although this solubilization does not change the total organic carbon content of the sludge, it does result in an increase in the BOD_5. The BOD_5 produced is of primary concern in the recycle of sidestreams. The solubilization of VSS and the resultant BOD_5 production for HT systems may be estimated with Equation 5-8:

$$\begin{aligned} VSS &= 0.1\ PS + 0.4\ WAS \qquad (5\text{-}8) \\ BOD_5 &= 0.07\ PS + 0.3\ WAS \end{aligned}$$

where,

- VSS = Volatile suspended solids solubilized, dry kg (lb)
- PS = Primary sludge, dry kg (lb)
- WAS = Waste Activated sludge, dry kg (lb)
- BOD_5 = 5-day biochemical oxygen demand produced by VSS solubilization, kg (lb)

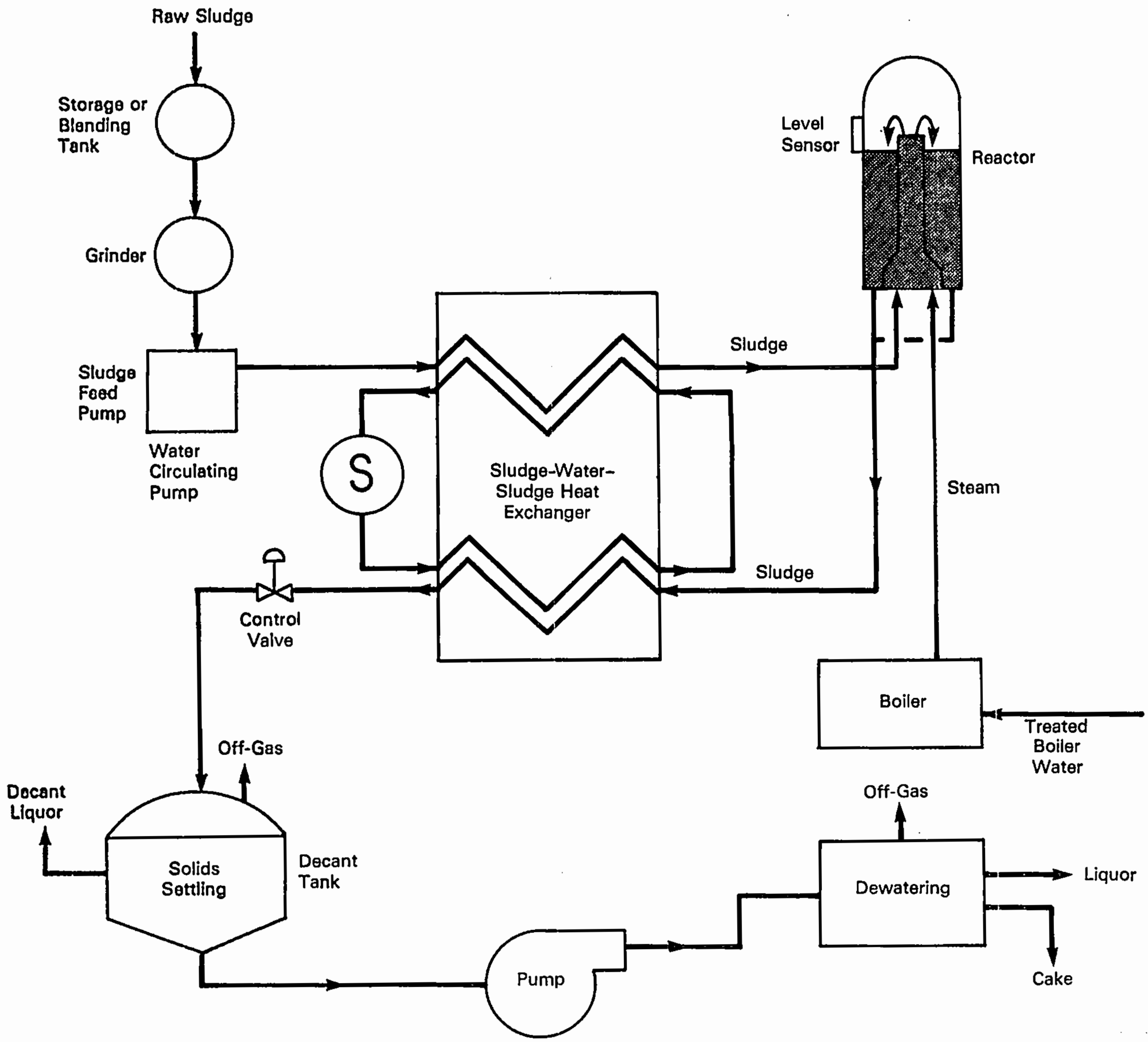
Using these rule-of-thumb procedures, 9.9 kg (22 lb) of VSS solubilization and 7.3 kg (16 lb) of BOD_5 are produced by heat treatment (HT) of 45 kg (100 lb) of a typical mixture of 60 percent primary and 40 percent waste activated sludge. In LPO systems, VSS solubilization and BOD_5 production are expected to be approximately the same.

Thermally conditioned sludge can be dewatered on vacuum filters, belt filter presses, recessed plate filter presses, centrifuges, or sand drying beds. The dewatered solids can then be incinerated or disposed of in a landfill or other land application method.

5.6.1 Heat Treatment

A schematic diagram of a typical HT system is shown in Figure 5-13. In this continuous process, raw sludge is ground to reduce particle size to less than 0.64 cm (0.25 in) and is then pumped through a heat

Figure 5-13. Heat treatment process flow diagram.

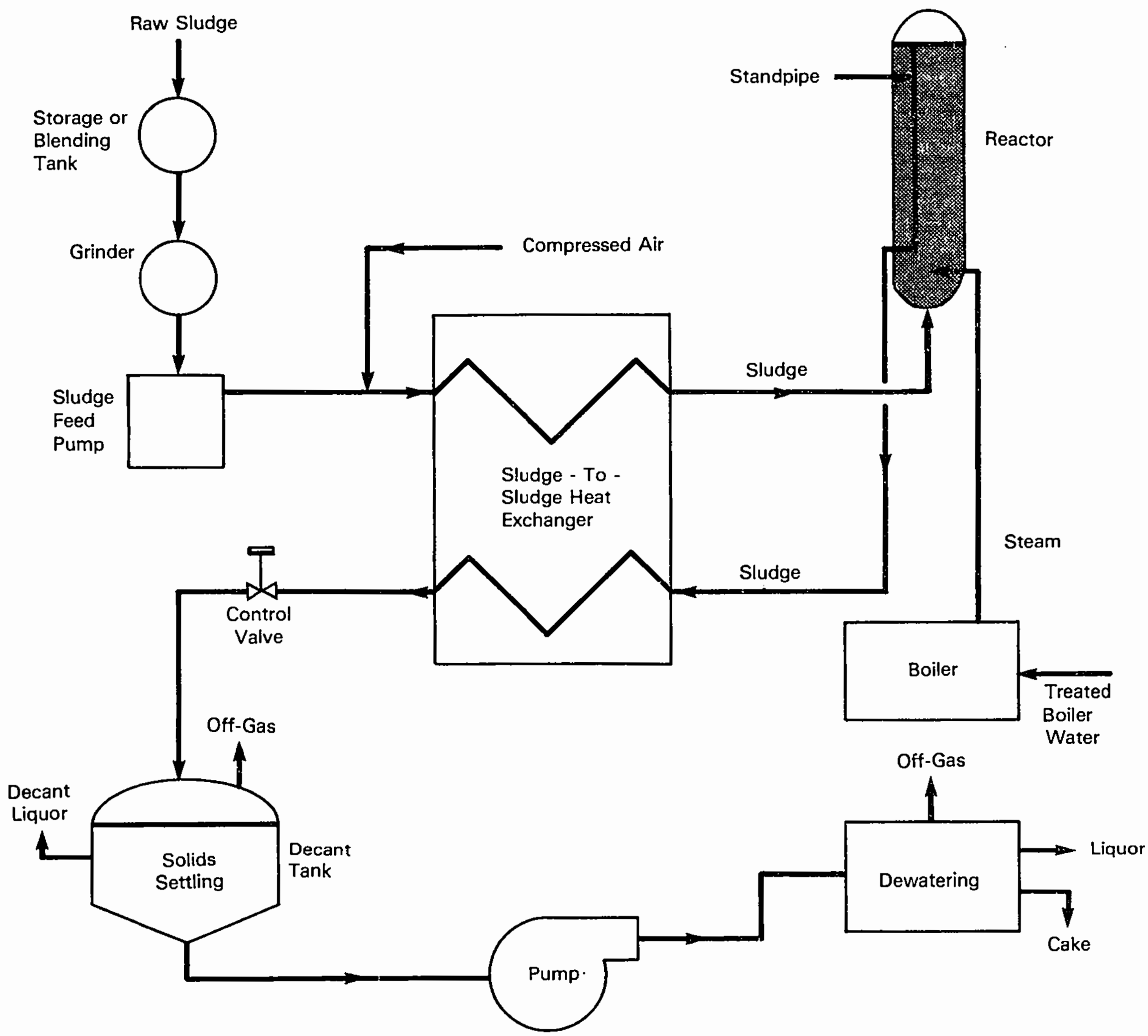


exchanger and into a reactor. Normal discharge pressure from the sludge feed pump is approximately 1,720 kPa (250 psi). In the heat exchanger, the temperature of the sludge is raised from ambient to between 149°C and 177°C (300°F and 350°F). The heated sludge exits the heat exchanger and enters a reactor feed standpipe, where steam is injected through a nozzle and the sludge is mixed turbulently. The steam and sludge proceed upward through the standpipe and enter the reactor at the top. The hot sludge is retained for a period of time in the reactor and is subsequently returned through the heat exchanger to be cooled at approximately 49°C (120°). From the discharge side of the heat exchanger, the conditioned sludge flows through a control valve, which controls reactor sludge level and pressure, and into a decant tank. The decant tank permits rapid settling and compaction of the sludge particles and the release of gas. The settled sludge is pumped to a dewatering device. Process off-gases can be treated by various odor control methods.

5.6.2 Low Pressure Oxidation

A schematic diagram of the LPO system is shown in Figure 5-14. Raw sludge is first passed through a grinder where particles are reduced to less than 0.64 cm (0.25 in). The ground sludge is then pumped at approximately 2,750 kPa (400 psi) through a heat exchanger followed by an LPO reactor. High pressure air from the system air compressor is introduced into the sludge flow upstream of the heat exchanger. The air improves heat transfer and converts sulfur

Figure 5-14. Low pressure oxidation process flow diagram.



products in the sludge to sulfate, slightly reducing odors from off-gases. The resulting turbulent flow of sludge and air proceeds through the heat exchanger where sludge is preheated by processed sludge returning from the LPO reactor. The sludge and air mixture enters the reactor at a temperature between 149°C and 160°C (300°F and 320°F). Steam is injected directly into the reactor to increase the sludge/air mixture temperature to between 166°C and 177°C (330°F and 350°F). The combined products rise slowly in the reactor and a slight heat of reaction or oxidation occurs, producing a small amount of heat. From the reactor midpoint to the reactor outlet, the sludge temperature increases approximately 10° due to the heat of reaction of the sludge, contributing to an overall temperature increase from the reactor inlet to reactor outlet of approximately 40°. Detention time or "cook time" in the reactor is based on the volume of the reactor and the height of the discharge pipe (standpipe or downcomer line). The detention time is controlled by the air, steam, and sludge flow rates to the reactor.

After leaving the LPO reactor, the partially oxidized product flows back through the heat exchanger and releases heat to the incoming sludge/air mixture. When the partially oxidized product reaches the control valve, the temperature ranges between 43°C and 54°C (110°F and 130°F). This valve controls the pressure in the reactor. From the valve, the thermally conditioned sludge and exhaust gases are released. The settled solids are then pumped to a dewatering device prior to final disposal. Process off-gases from the LPO system also can be treated by various odor control methods.

5.6.3 Economic Considerations

The increase in the cost of natural gas and fuel oil since the early 1970s has significantly changed the economic feasibility of new thermal conditioning systems for small plants. Larger installations, greater than 0.44 m^3/sec (10 mgd), that use dewatering and incineration with energy recovery may determine that the addition of a thermal conditioning step would be an economic asset.

Several factors must be considered regarding the cost effectiveness of a thermal conditioning system as a function of plant size.

- Present-day energy costs dictate some form of heat recovery to make the thermal conditioning process competitive with other conditioning processes.
- Thermal conditoning systems require well trained and skilled supervisors and operators to optimize the operation and maintenance of the systems.
- Both types of systems should be supported with a complete inventory of spare parts to reduce excessive downtime. Also, they require a thorough preventive maintenance program.
- The unit capital cost of thermal conditioning systems is in the range of $385 to $550/Mg ($350 to $500/ton) of annual sludge production, when processing over 9,090 dry Mg (10,000 tons) per year due to use of multiple treatment units and standby units rather than larger sized individual units. At lower loading rates, processing costs increase significantly, and the comparatively high cost of support systems (such as boilers, air compressors, and decant tanks) makes HT/LPO systems more costly to build than other sludge conditioning facilities.

5.6.4 Advantages and Disadvantages of HT/LPO Conditioning

Previous literature on HT/LPO provides a summary of the advantages and disadvantages of using these processes to condition wastewater sludges.

Advantages cited include:

- Except for straight waste activated sludge, the process produces a sludge with excellent dewatering characteristics. Cake solids concentrations of 30 to 50 percent can be obtained with conventional dewatering equipment.
- The processed sludge does not normally require chemical conditioning to dewater well on mechanical equipment.
- The process stabilizes the sludge and destroys all living organisms including pathogens.
- The process provides a sludge with a heating value of 26,000 to 30,000 KJ/kg (11,000-13,000 Btu/lb) of volatile solids, suitable for incineration or anaerobic digestion with energy recovery.
- The process is suitable for many types of sludges that cannot be stabilized biologically because of the presence of toxic materials.
- The process is effective on feed sludges with a broad range of characteristics and is relatively insensitive to changes in sludge characteristics.
- Continuous operation is not required as with incineration, since the system can easily be placed on standby.

Disadvantages cited include:

- The process has high capital costs due to mechanical complexity and the use of corrosion-resistant materials, such as stainless steel, in the heat exchangers.
- The process requires careful supervision, skilled operators, and a good preventive maintenance program.
- The process produces a malodorous gas stream that must be collected and treated before release.
- The process produces dark colored sidestreams with high concentrations of organics and ammonia nitrogen.
- Scale formation in heat exchangers, pipes, and the reactor requires cleaning by difficult and/or hazardous procedures.
- Subsequent centrifugal dewatering may require continuous or intermittent polymer dosage to control recycle of fine particles.
- The daily sludge throughput of the process cannot be adjusted by a significant amount without incurring high energy and/or labor costs.

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Chapter 6
Air Drying Processes

6.1 Introduction

As used in this manual, "air drying" refers to those dewatering techniques by which the moisture is removed by natural evaporation and gravity or induced drainage. There may be some mechanical assistance, such as turning and mixing the sludge on paved beds, or some vacuum assistance but the movement of water is controlled by natural forces. The air drying processes described in this chapter range from the oldest dewatering concept (sand beds) to some recently developed techniques. They include:

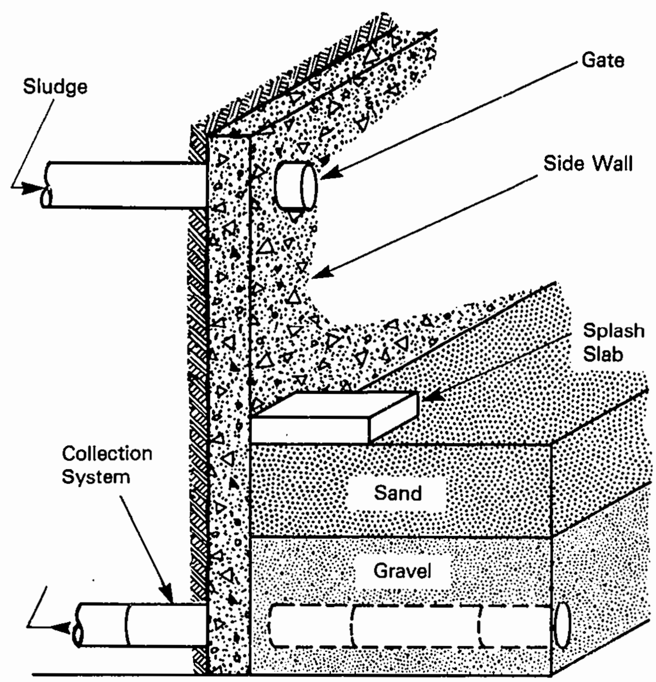
- Sand beds
- Freeze assisted sand bed dewatering
- Vacuum assisted beds
- Wedgewire beds
- Sludge lagoons
- Paved beds
- Other innovative processes.

Air drying processes are less complex, easier to operate, and require less operational energy than mechanical dewatering systems. They do, however, require a larger land area and more labor, primarily for sludge cake removal. The combination of all of these factors suggests that air drying processes are especially well suited for small to moderate sized communities with design wastewater flows less than 7,500 m^3/d (2 mgd). Air drying processes are technically feasible at greater flows but the need for the larger land area restricts the economic feasibility in some locations. Air drying processes should be given strong consideration for all small to moderate sized communities, and for larger facilities in arid and semi-arid climates when land is available. Sand beds, freeze dewatering, and reed beds can easily produce a sludge cake with 25 to 40 percent solids and can exceed 60 percent solids with additional drying time. These three processes can produce a drier sludge than any of the mechanical dewatering devices discussed in Chapter 7.

6.2 Sand Beds

Sand beds have been used successfully for sludge dewatering since wastewater treatment became a recognized technology early in this century. In 1987, they are still the most frequently used technique for sludge dewatering in the United States. Dewatering on the sand bed occurs through gravity drainage of free water followed by evaporation to the desired solids concentration level. Figure 6-1 illustrates details of a typical sand bed. (In areas of high precipitation, covered sand beds have been used.)

Figure 6-1. Sand bed details.



Sidewalls can be constructed of reinforced concrete as shown in Figure 6-1, or treated timber planks or concrete planks. The plank type construction has the advantage of allowing adjustment of the total depth of the bed, an important feature for the freeze dewatering process discussed in Section 6.3.

6.2.1 Design Considerations

The critical design parameter is the surface area of the sand bed required to attain the necessary drainage and evaporation in the specified time. Most of the sand beds in current use were designed with per capita loading criteria that were developed empirically in the early 1900s (1,2). Some State agencies still specify drying bed criteria on the basis of per capita loading; so it is necessary to obtain these values prior to any specific project design. These per capital loading criteria are still valid if all of the original conditions are incorporated, but this is not always the case in modern systems. The preferred approach is to base design on the mass loading of solids. Currently accepted loading criteria are presented in Table 6-1 (1). The values for digested primary plus waste activated sludge (WAS) for uncovered beds range from 60 to 100 kg/m^2/yr (12 to 20 lb/sq ft/yr). These values can be increased to 85 to 140 kg/m^2/yr (17 to 28 lb/sq ft/yr) for covered beds, and a significant further increase is possible if coagulants are used to condition the sludge prior to application on the bed (1). The upper end of these ranges applies for warm dry climates and to sludges which drain readily.

Table 6-1. Loading Criteria for Anaerobically Digested, Non-Conditioned Sludge on Uncovered Sand Beds

Sludge Type	Mass Loading
	kg/m^2/yr
Primary	120-200
Primary	100-160
Primary plus low-rate TF	100-160
Primary plus WAS	60-100

kg/m^2/yr x 0.2048 = lb/sq ft/yr.

A sand bed's performance depends on:

- The required solids concentration in the dewatered sludge
- The solids concentration in the applied sludge
- The type of sludge to be applied (e.g., stabilized, thickened, conditioned)
- The drainage and evaporation rates.

The required solids concentration depends on the technical or regulatory requirements for final sludge disposal or utilization (see Chapter 2 for discussion). If no special requirements apply, the sludge cake is typically "liftable" at about 25 to 30 percent solids and can be removed from the bed without excessive sand loss.

Water is removed from the sludge through gravity drainage and evaporation. The amount of water that can be removed by drainage is strongly influenced by the type of sludge applied. Drainage might account for 25 percent of water removal for some anaerobically digested primary plus waste activated sludges, and 75 percent or more for well conditioned sludges. Significant drainage is typically complete within 3 to 5 days. However, the unit drainage rate is less critical than the total percentage of water removed (percent total solids).

The rate of evaporation is a function of local climatic conditions and the sludge surface characteristics. Seasonal evaporation rates can be obtained from local pan or lake evaporation values. Since the crust which forms on the sludge surface inhibits evaporation, the pan evaporation values must be adjusted when designing the sand bed. An adjustment factor of 0.6 was experimentally derived (3) (see Section 6.7 for further discussion). Once "cracking" occurs (see Figure 6-9), the evaporation rate should approach the pan value due to the additional sludge surfaces exposed.

Design equations which relate the initial and final solids concentration and the amount of water lost to drainage and evaporation are presented below. These design equations provide a rational method of determining the design mass loading on a sand bed and other critical operational parameters.

The drying time for a single application is given by:

$$t_d = \frac{(y_0)(1 - s_0/s_f)(1 - D)}{(k_e)(E_v)} \qquad (6-1)$$

where,

t_d = dewatering time for a single application, mo

y_0 = initial depth of applied sludge layer, cm (in)

s_0 = initial dry solids concentration required for dewatered sludge, percent

s_f = final dry solids concentration required for dewatered sludge, percent

D = fraction of water removed by drainage, percent as decimal

E_v = average pan evaporation during time t_d, cm/mo (in/mo)

k_e = reduction factor for sludge evaporation vs. a free water surface, percent as decimal
= 0.6 (a pilot-test is recommended to determine this value)

The number of applications during the operating season is given by:

$$N = \frac{n_v}{t_d} = \frac{(n_v)(E_{vn})(k_e)}{(y_0)(1 - s_0/s_f)(1 - D)} \qquad (6-2)$$

where,

N = number of sludge applications

n_v = length of operating season, or an increment if evaporation is significantly different, mo

E_{vn} = average pan evaporation during period n_v, cm/mo (in/mo)

The design solids loading is given by:

$$L = (C)(s_0)\frac{(n_v)(E_{vn})(k_e)}{(1 - s_0/s_f)(1 - D)} \qquad (6-3)$$

where,

L = solids loading during period n_v, kg/m^2 (lb/ft^2)

C = Conversion factor (assumes specific gravity of sludge = 1.04)
= 10.4 (metric units)
= 5.41 (US units)

The annual solids loading is determined by summation of the results of equation 6-3 for the different operational periods selected, over a full 12-month annual cycle. To obtain a first approximation, assume nv = 12 and use the average annual pan evaporation for E_{vn}, if the beds are to be operated on a year-round basis.

The final depth of the dewatered sludge cake is given by:

$$y_f = (y_0)\frac{s_0}{s_f} \qquad (6-4)$$

where,

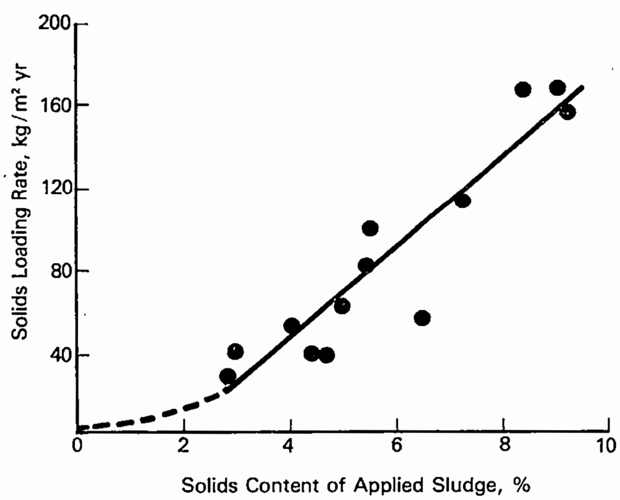
y_f = final depth of dewatered sludge cake, cm (in)

Thin layers of sludge will dry faster than a thick layer, but (as defined by Equation 6-3) the annual solids loading is independent of the depth of the individual layers applied. Using too thin a layer has several disadvantages, including more frequent operation and maintenance (some dried sludge must be removed before the next application), greater sand loss from the bed, and increased costs in general. A 20 cm (8 in) layer, as recommended in many design texts, at 3 percent solids would produce a dewatered cake at 30 percent solids only 2 cm (0.8 in) deep. Such a depth is too thin for most mechanical removal techniques and could result in excessive sand loss.

To keep operation and maintenance costs as low as possible, the design goal is to achieve the maximum possible solids loading with the minimum number of application and removal cycles. Repeated calculations with Equations 6-1 through 6-4 will converge on the most effective combination of initial solids concentration and layer depth for a particular project. Final optimization of the layer depth is only possible with operational experience.

The annual solids loading depends on the solids concentration in the applied sludge, as shown by Equation 6-3. This relationship is demonstrated by Figure 6-2 using actual data from 13 operational systems in Pennsylvania, Ohio, New York, California, Texas, Illinois, and North Carolina (4,5).

Figure 6-2 Solids loading rates for sand beds vs. solids content of applied sludge.



An increase in solids content from 2 to 4 percent, for example, would approximately double the loading rate and could reduce the required bed area by one half. This relationship demonstrates the potential advantage of thickening or preconditioning the sludge prior to application on the bed. However, increasing the solids content beyond 8 percent is not recommended since the sludge will not flow and distribute uniformly on the bed beyond this level. At existing facilities where expansion may not be possible, the use of prior sludge thickening might allow more effective use of the present bed area.

Typically, the total bed area is subdivided into multiple cells. It is convenient to size the cells so that one or two can contain the total volume of sludge from a scheduled digester withdrawal. The width of the bed depends on the removal method. Small to moderate sized facilities with hand or semi-mechanical removal are about 6 m (20 ft) wide. Greater widths are used with mechanical removal methods; each cell must contain an entry ramp and possibly paved runway slabs for equipment. Sand beds as long as 30 to 60 m (100 to 200 ft) have been successfully used with the more dilute sludges. Uniform sludge distribution on the bed can be difficult, particularly when polymers are used for conditioning. In these cases, the bed length should not exceed 15 to 25 m (50 to 75 ft) and/or multiple distribution points should be incorporated into the design.

6.2.2 Structural Elements

Sludge can be applied to each cell with a valved pipe (plug valves) or from an open channel with gate controls along the perimeter of the bed. The open channel is easier to clean, but is more difficult to operate in cold weather. The valves in a pipe network should be protected from freezing in cold climates, since the adjacent pipe will not always drain completely. A splash block on the bed at every sludge entry point minimizes erosion of the sand.

A minimum sand depth of 30 cm (12 in) is recommended. In some cases, depths of up to 46 cm (18 in) can be used to extend the life of the bed. Since sand is unavoidably removed every time sludge is taken from the bed, new sand must eventually be installed. Preferred characteristics for the sand are:

- Clean, hard particles - no clay, silt, or organic matter
- Effective size 0.3 to 0.75 mm (0.01 to 0.03 in)
- Uniformity coefficient <3.5.

The gravel layer is usually 20 to 46 cm deep (8 to 18 in), with gravel sizes ranging from 3 to 25 mm (0.1 to 1.0 in). With mechanical sludge removal, greater depth of gravel is needed to structurally protect the underdrain network. A thinner layer of coarser stone, overlain by a suitable permeable geotextile membrane, can be used with hand removal or mechanical removal with very light equipment.

Underdrains are usually plastic pipe or clay tile laid with open joints. The main underdrain pipes should be at least 10 cm (4 in) in diameter and should be laid with a slope of at least 1 percent to insure drainage. Spacing of these main underdrains ranges from 2.5 to 6 m (8 to 20 ft) depending on the type of sludge removal planned. Lateral drain pipe branches connected to the main drains should be on about 2.5 m (8 ft) centers.

Covered beds have often been used in northern areas to extend the otherwise seasonal dewatering operations. Standard designs for glass or plastic enclosures are available. The freeze dewatering technique described in Section 6.3 allows the use of uncovered beds throughout the winter in cold climates. The covered beds might still be desirable in the marginal zones shown on Figure 6-3 where the potential for freeze dewatering is limited.

6.2.3 Performance Expectations

The drainage rate during the initial dewatering phase depends on the type of wastewater and sludge treatments used, on the concentration of solids, and the depth of sludge applied. Well stabilized aerobically digested sludges will tend to drain more completely than anaerobically digested sludges, for example, and most raw sludges will drain more readily than digested sludges. Drying of raw sludges, however, is likely to result in malodorous conditions in all but dry warm climates.

The drainage phase is usually measured in terms of hours or a few days. The evaporative stage lasts as long as is necessary for the sludge to reach the desired solids concentration as defined by Equation 6-1. The most significant volume reduction occurs up to about 30 to 40 percent solids. Further drying beyond that point achieves little additional volume reduction but may still be required by regulatory authorities. Sludge at about 30 percent solids can be removed from the bed with minimal loss of sand.

The water collected in the underdrainage network is returned to the wastewater treatment facility. Characteristics of this liquid will depend on the type of treatment process used, and may be similar to that reported in Table 6-2.

Table 6-2. Sludge Filtrate Characteristics After Freeze Thawing

		Digested Sludge	
Parameter	Raw WAS	Aerobic	Anaerobic
BOD, mg/l (dissolved)	706	722	1,012
COD, mg/l	1,585	1,815	3,325
SS, mg/l	14	17	18
Total P, mg/l	28	46	80

6.2.4 Operation and Maintenance

The optimum depth of sludge to apply will be determined with experience; in general this depth may range from 20 to 45 cm (8 to 18 in). Any chemical conditioners should be added continuously during the pumping operation, at points in the system that will insure proper mixing. Multiple dosage points for polymers should be constructed into the system. These dosage points, at a minimum, should be located ahead of the pump suction, at the pump discharge, and ahead of the discharge point to the bed. It may not be necessary to use all dosage points but the multiple array will allow optimization after operation commences.

Bed maintenance involves the periodic replacement of sand lost during sludge removal, leveling and scarification of the sand surface prior to dosing, and removal of vegetation. Odors should not be a problem with well stabilized sludges. To control odors, calcium hypochlorite, potassium permanganate or ferrous chloride can be added to the sludge during application to the bed.

The time required for O & M activities is a direct function of the size of the system and the number of operational cycles used in a year. Very small systems (<100 m^2, at 100 kg/m^2/yr) might require about 4 hr/yr m^2 (0.4 hr/yr/ft^2), while larger systems (>4,000 m^2, at 100 kg/m^2/yr) need less than 0.5 hr/yr/m^2 due to the economies of scale and increased mechanization.

6.2.5 System Upgrading

At many locations the original sand bed capacity may become inadequate. Causes include increased system flow and sludge production or the use of chemicals in wastewater treatment that can significantly increase the mass and volume of sludge to be handled, as well as adversely change the drainage characteristics of the sludge. Alum, for example is often used for phosphorus removal, but the resulting sludge does not readily drain on sand beds. A similar situation often occurs with iron salts. Some of these concerns are reflected in Table 6-1 with the larger sand bed area allowance for "Primary plus chemicals." Similar allowance must be made for "Primary plus waste activated sludge" when aluminum or iron salts are present. Typically, a plant operating at capacity that decides to add aluminum or iron salts for phosphorus removal may need to take action in one or more of several ways. These include constructing additional drying beds, adding polyelectrolytes to facilitate dewatering, modifying sludge application procedures, lengthening sludge drying times, and replacing sand beds with some type of mechanical dewatering device. Alternatives that must be considered include the use of polymer conditioning alone or combined with sludge thickening. Either approach may allow the more effective use of existing sand beds, or in new designs allow the use of sand beds where land limitations exist.

In the typical case with anaerobically digested sludge, about 25 to 60 percent of the total water applied is removed by drainage. The use of an appropriate polymer accelerates the particle agglomeration and increases the amount of water which can be drained. Polymer conditioning reduces the amount of water to be evaporated and thereby allows a significant increase in solids loading.

The use of polymers for conditioning is discussed in detail in Chapter 5. The selection of polymer type and optimum dose is based upon the results of tests with the sludge in question. These tests are usually series of Buchner funnel or capillary suction time (CST) tests. Typical costs range from $3.00 to $11.00 per ton of dry sludge solids (see Chapter 8 for case studies). Sludge characteristics can change and it is often necessary to vary the dose or the polymer type occasionally. Sludges that drain poorly are the best candidates for polymer conditioning. Polymers are also a benefit with thin sludges having high concentrations of fine particles. Sludges of this type tend to penetrate beyond the sand surface and eventually plug the bed. Polymers increase drainage rate and reduce penetration of sludge particles.

Even if polymers have been used, sludges that drain poorly will usually still require additional drying time in the evaporative stage to reach the desired solids concentrations. The use of polymers can increase bed capacity by inducing greater drainage and thereby reducing the time required for evaporative drying, but the evaporation rate is unaffected (6,7).

Problems can occur if either mixing or distribution are inadequate, which can cause localized deposits and clumps of sludge which are then slow to drain. To insure the effective use of the entire bed area when using polymers, multiple inlet points are recommended.

6.2.6 Costs

The capital cost for sand beds is strongly influenced by the cost of land at the project site. Other major factors include the containing walls and bed bottom, the application and drainage piping, and any sludge removal equipment. The major O & M costs are labor, fuel for equipment, periodic sand replacement, and conditioning chemicals (if used). Section 5-8 in Reference 8 can be used to estimate capital and O & M costs for sand beds. Section 6-5 in the same reference should also be used if polymer conditioning is planned. The design example in Appendix A demonstrates the use of these procedures.

6.3 Freeze Assisted Sand Bed Dewatering

Freezing and then thawing a sludge will convert a material with a jelly-like consistency to a granular type material that drains readily. Solids concentrations exceeding 20 percent will be realized as soon as the material is thawed and 50 to 70 percent can be achieved with minimal additional drying time (9). The effects of sludge freezing have been recognized for over 50 years but until recently a generally applicable design procedure was not available (9).

Freezing will work with any type of sludge, at any solids concentration, but is particularly effective with chemical and biochemical sludges that do not drain readily. Energy costs for artificial freeze-thawing are prohibitive so the concept must depend on natural freezing to be cost effective. As shown in Figure 6-3 the feasible area can include most of the northern half of the United States. The recently developed design procedure will allow the year-round use of new and existing uncovered sand beds in colder climates.

6.3.1 Design Considerations

The freeze-dewatering effects will be realized regardless of the initial sludge concentration or the degree of stabilization, but the cost effectiveness of the operation will be influenced by both factors. A very dilute sludge will increase costs by requiring more area for freezing beds; thickened sludge in the range of 3 to 7 percent solids works well. The use of stabilization for wastewater sludges is recommended to avoid odor complaints during thawing and drying and to meet regulatory requirements for final disposal.

The design of a freeze dewatering system must be based on worst case conditions to insure successful performance at all times. If sludge freezing is to be a reliable expectation every year, the design must be based on the warmest winter during the period of concern (usually 20 years) and on a layer thickness which will freeze within a reasonable time if freeze-thaw cycles occur during the winter. It is essential for the layer to freeze completely to achieve the dewatering benefits. In many locations a large single layer may never freeze completely to the bottom, with only the upper portion going through alternating freezing and thawing cycles. Recent research (9) has indicated that an 8-cm (3-in) deep layer of sludge is practical for most locations in moderately cold climates. A thicker layer is feasible in colder climates; facilities in Duluth, MN, for example, successfully freeze water treatment sludges in 23 cm (9 in) layers and a 46-cm (18-in) layer has been used in Fairbanks, AK. An 8-cm (3-in) layer should be assumed for feasibility assessment and preliminary design. A larger increment may then be justified by a detailed evaluation during final design. The freezing or thawing of a sludge layer can be described with Equation 6-5:

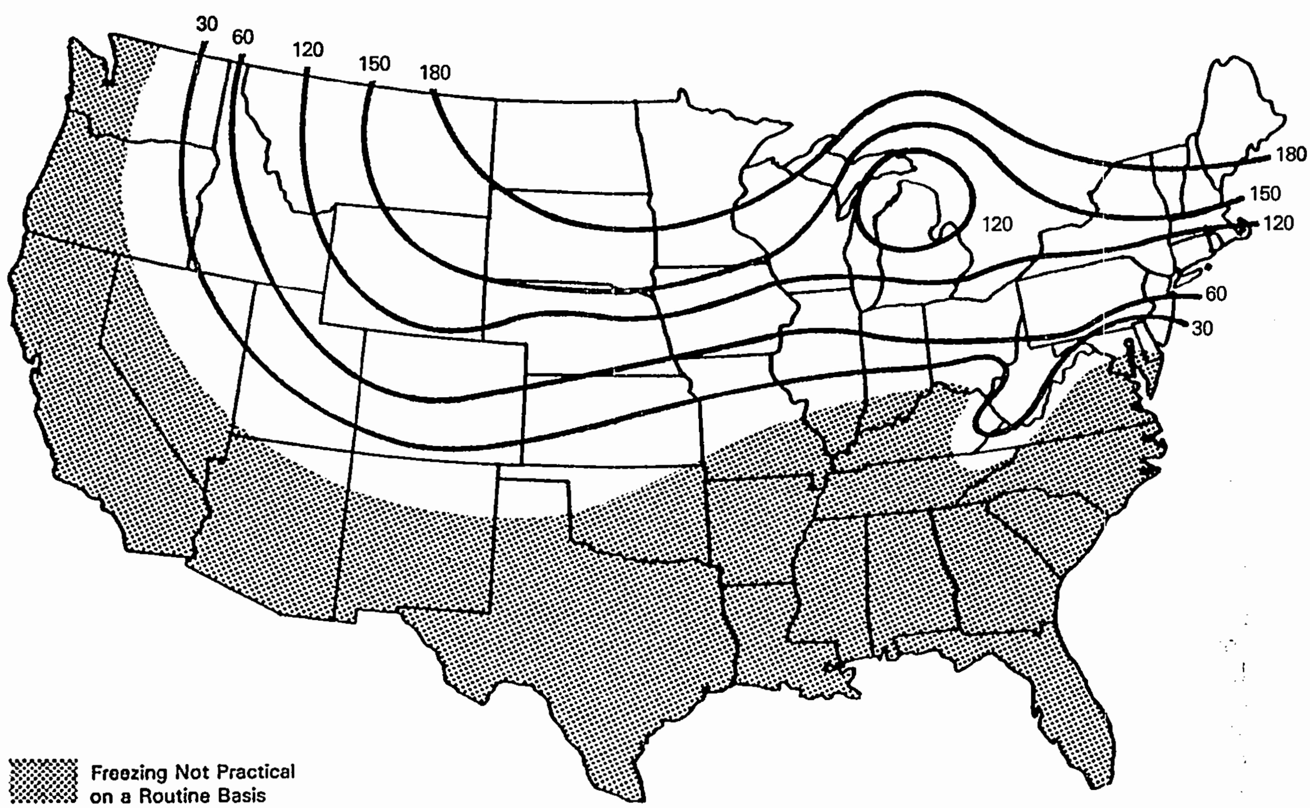
$$Y = (m)(\Delta T \bullet t)^{1/2} \qquad (6-5)$$

where,

Y = depth of freezing or thawing, cm (in)

m = proportionality coefficient, dependent on thermal conductivity, latent heat of fusion and density of the material being frozen, $cm(°C \bullet d)^{-1/2}$ $[in(°F \bullet d)^{-1/2}]$

Figure 6-3. Potential depth of sludge (cm) that could be frozen, if applied in 8-cm layers (13).



$\Delta T \bullet t$ = freezing or thawing index, °C•d (°F•d)

ΔT = difference between average ambient air temperature and freezing temperature, °C (°F)

t = time period of concern, days

The proportionality coefficient was experimentally determined with wastewater sludges and should be valid over the range 0 to 7 percent solids. The freezing or thawing index is an environmental characteristic for a particular location. The values are sometimes published but can also be determined from weather records. For example:

Average daily air temperatures: 0°C, -3°C, -7°C, -4°C

Time period t = 4 d

Average temperature during period = -3.5°C

Freezing index for the period
= [0°C -(-3.5°C)](4)
= 14°C•d.

Since a layer thickness of 8 cm was suggested for preliminary design, it is possible to rearrange Equation 6-1 and solve for the time to freeze the design layer using Equation 6-6:

$$\sum \Delta T \bullet t = (Y/m)^2 \qquad (6-6)$$

with Y = 8 cm and m = 2.04 cm (°C•d)$^{-1/2}$, equation 6-6 becomes

$$\sum \Delta T \bullet t = 15.38\ °C \bullet d$$

This form is used with the local weather records to determine how many 8-cm (3-in) layers can be frozen during each winter of the study period. The year with the smallest number of layers is then the control year for design.

It can be assumed, for example, that the first layer is applied in late fall, and Equation 6-6 is then used to determine the number of days required to freeze the layer under the average temperature conditions indicated in the records. Either the intensity or the duration of the low temperature must be sufficient to freeze the layer in a continuous period. The calculations are repeated for the entire winter season with a 1-day allowance for each sludge application and cooling, and due account taken of any thaw periods during the winter. The next layer is not applied until calculations show that the previous layer has frozen completely. This procedure can be easily programmed for rapid calculations with a small computer or desk-top calculator.

6.3.1.1 Preliminary Feasibility Assessment

A rapid method for preliminary assessment and design relates the potential depth of sludge which may be frozen in the "design" year to the maximum depth of frost penetration at a particular location. There is a high correlation between the two factors since they both depend on the same environmental conditions. The maximum depth of frost penetration for an area can be found in local records or other published sources (10). The relationship between the two factors is defined by Equation 6-7:

$$\sum Y = 1.76\,F_p - 101 \qquad (6-7)$$

where,

ΣY = total depth of sludge which could be frozen in 8-cm layers during the "design" year, cm

F_p = maximum depth of frost penetration into the soil for the location, cm

In US units, with Y and F_p in inches, the equation is:

$$\sum Y = 1.76\,F_p - 38$$

Equation 6-7 is the basis for the map shown in Figure 6-3. The map and Equation 6-7 are only valid for preliminary estimates. Detailed weather records and Equation 6-6 should be used for final design. Although very effective, freeze dewatering is a seasonal process. Except in very cold climates it is not economical to store sludge in the warm months and depend only on winter freezing for dewatering. In most parts of the United States it will be more cost effective to combine winter freezing with polymer assisted summer dewatering on the same beds. This combination of techniques would eliminate the need for large scale sludge storage and reduce the total number of beds required. Figure 6-4 demonstrates the application of Equation 6-6. The arrows and circles on the diagram represent the predicted times for sludge application and layer freezing at Duluth, MN. The predicted total depth is the same as was actually observed at the treatment facility (11). This design approach has also been independently verified with full-scale tests in Sweden (12).

6.3.2 Structural Aspects

The basic facility is essentially the same as for conventional sand drying beds. The major design difference is increased freeboard to contain the design depth of frozen sludge. The maximum potential for freezing will occur when the sludge is exposed to the extreme weather conditions, so covering the bed, using windbreaks, or applying the

Figure 6-4. **Predicted vs. measured sludge freezing at Duluth, MN.**

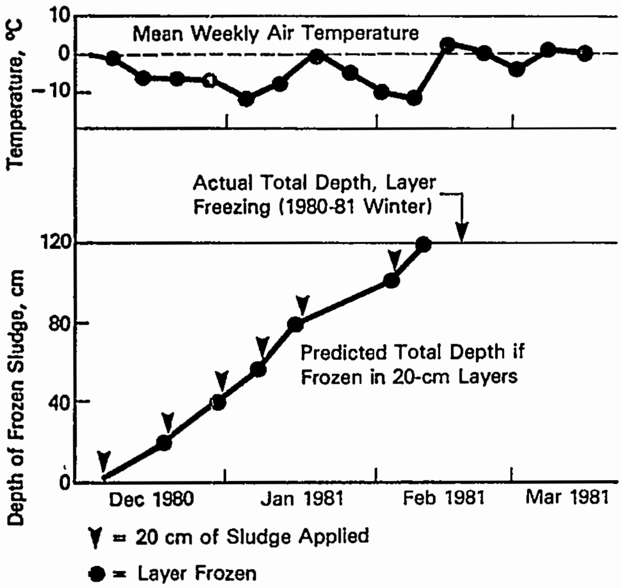
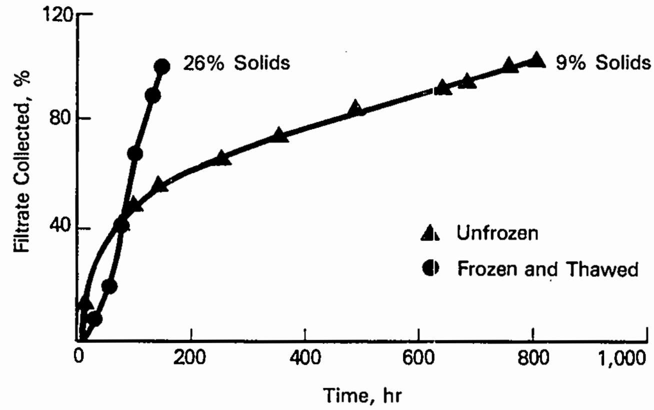


Figure 6-5. **Effect of freeze thawing on the drainage rate for anaerobically digested sludge.**



sludge in a deep trench will only reduce the freezing rates. Trenches are used at the system in Duluth, MN (11), for sludge storage during the warm months. Supernatant is decanted prior to the onset of winter. As soon as a surface layer of sludge is frozen, a hole is drilled in the ice and sludge is pumped up to freeze in layers on top of the ice.

An effective way to provide the necessary freeboard, and still allow exposure of the sludge to winter conditions, is the use of concrete or timber planks to increase the sidewall depth of the bed as the winter progresses. The sludge feed system must be designed to apply each layer on top of the previously frozen material. A hydrant and hose combination is one possibility.

6.3.3 Performance Expectations

Freezing a sludge changes both the structure of the sludge water mixture and the characteristics of the solid particles. In effect, the solid matter tends to be compressed into large discrete conglomerates surrounded by frozen water. When thawing commences, drainage occurs instantaneously through the large pores and channels created by the frozen water. Cracks in the frozen mass also act as conduits to carry off the melt water.

Experience in a number of locations (9,12,13) has shown that the solids concentration will approach 25 percent as soon as the frozen mass is completely thawed, due to the very rapid drainage. Figure 6-5 shows the drainage rate of a frozen sludge after thawing as compared to the drainage rate of the same sludge without freezing.

The time required for thawing can be estimated with Equation 6-6, using an "m" value of 3.78 cm $(°C•d)^{1/2}$ (1.11 in$(°F•d)^{1/2}$) (13). In this case the coolest expected spring/summer temperatures should be used for design purposes, and the depth to be used in the equation is the total depth of frozen material, not the individual layers. In extremely cold climates it would be possible to freeze more sludge, in thin layers, than could be thawed in the very short summer; this is unlikely to occur anywhere in the continental U.S. However, the time to thaw should be calculated for all locations above the 150 cm line on Figure 6-3 to insure that the frozen sludge will thaw in time for the bed to be used in the conventional manner during the spring and summer.

The maximum potential response during both the freezing and thawing portions of the cycle can be obtained by exposing the sludge on open uncovered beds. Section 6.3.4 provides operational guidance for rain or snow conditions during the freezing, thawing, or drying phases.

The drainage of water during thawing may occur at a faster rate, and will produce a greater volume when compared to applying the same unconditioned sludge to a conventional sand drying bed. Typical characteristics of this drained liquid are given in Table 6-2 (14). Also, the freezing and thawing process will not improve the pathogen kill in the sludge. Freezing conditions preserve rather than destroy most pathogens.

6.3.4 Operation and Maintenance

The critical operational requirement is to ensure complete freezing of the sludge layer before the next is applied. Hand probing with a small pick or axe is the easiest way to make this determination at small facilities. Remote temperature sensing devices, such as thermocouples, can be used but will not directly indicate when freezing commences or exactly when the mass is completely frozen since the temperature will remain at 0°C until all of the latent heat is drawn off. The operator should maintain records of average daily temperature and other weather conditions during the freezing, thawing, and drying phases. With this data and some experience, the operator can quickly develop site specific criteria for sludge applications and removal.

Rainfall during the final drying phase seems to have few lasting effects. Rainfall during the freezing stages may thaw some of the previously frozen material. If time permits, the thawed and drained material can be removed; otherwise the next sludge application should not be made until any melt water has drained or refrozen. A light snowfall, <5 cm (2 in) just prior to or during the freezing stage is not a concern since the mass of water involved is small and the snow will help in the initial sludge cooling. A heavy snowfall will act as an insulating barrier and retard the freezing rate. Snow layers greater than 5-8 cm (2-3 in) should be removed from the bed with a snow blower or front-end loader prior to the next sludge application to ensure maximum exposure of the sludge.

In most cases it will be more cost effective to combine freeze thaw dewatering with polymer assisted dewatering and/or thickening in the warm months on the same drying beds. In order to optimize bed use, the operator should only apply that quantity of sludge for freezing that can be removed from the bed as dried cake by mid-May of each year. The system design will provide a conservative procedure based on worst case conditions. The operator can make the necessary adjustments depending on the weather conditions during each winter.

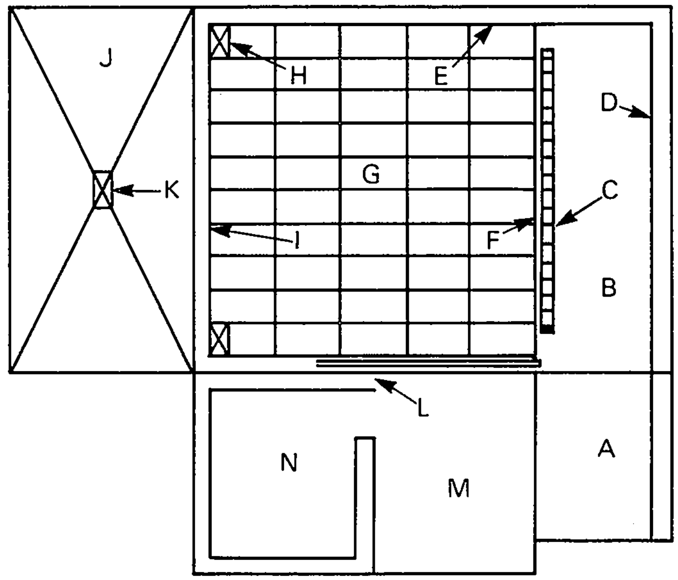
6.3.5 Costs

The capital costs for freeze thaw dewatering beds are essentially the same as described in Section 6.2.6 for uncovered, conventional sand drying beds. Additional capital costs may be required for the extra timber or concrete planks and for thermal protection of any exposed sludge piping, since the system must operate throughout the winter. Operation and maintenance costs should be about the same as for conventional sand drying beds (see Section 6.2.6).

6.4 Vacuum Assisted Drying Beds (VADB)

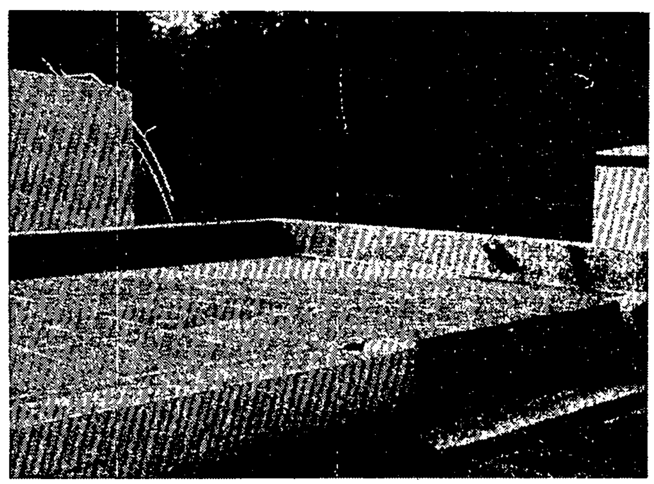
This dewatering technology applies a vacuum to the underside of rigid, porous media plates on which chemically conditioned sludge has been placed. The vacuum draws free water through the plate and essentially all of the sludge solids are retained on top, forming a cake of fairly uniform thickness. Figure 6-6 is a plan view of a typical single bed system showing the necessary or desirable support components, Figure 6-7 is a typical outdoor facility, and Figure 6-8 is a cross section view of a typical epoxy-bonded media plate.

Figure 6-6. Plan view of a vacuum assisted drying bed system.



A. Entrance Ramp
B. Off-Bed Level Area
C. Area Drain
D. Curbing
E. Sludge Distribution Piping
F. Bed Closure System
G. Media Plates
H. Corner Drain
I. Bed Containment Wall
J. Truck Loading Area
K. Area Drain
L. Wash Water Supply
M. Sludge Feed Inventory Tank (below grade, seldom required)
N. Control Building with
 - Sludge Feed Pumps
 - Polymer System
 - Vacuum Pumps
 - Control Panel
 - Filtrate Receiver/Pumps (below grade)

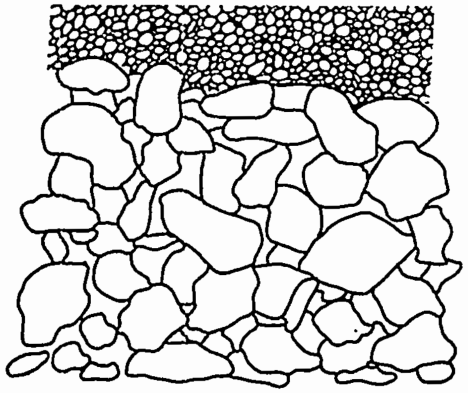
Figure 6-7. Typical outdoor vacuum assisted drying bed.



6.4.1 Design Considerations

The basis for design is the average annual sludge (dry solids) production rate and the number of cycles which can be conveniently carried out in a typical

Figure 6-8 Cross section of a typical vacuum bed media plate.



work week. To insure reliability, the design may require an increase in the size or number of beds. A two-bed system should be the minimum standard. If the sludge production rate exceeds one dry ton of solids per day, a three-bed system is recommended.

A properly sized three-bed system, using a 24-hour total cycle time, would utilize two of the beds for dewatering each operating day with the third bed idle. Each bed in such a system should be sized to dewater, at a minimum, 70 percent of the average daily sludge mass drawn from the treatment system. This design will allow the dewatering system to be operated a maximum of 5 days per week and still provide for the dewatering of 7 days' accumulation of sludge. In the event the applied sludge did not properly dewater within the allowed 24 hours, the third bed could be used during the succeeding days to dewater one-half of the daily sludge production, or the other two beds could be used over the weekend to get the system back on schedule.

A solids loading of about 10 $kg/m^2/cycle$ (2 $lb/ft^2/cycle$) has been found acceptable. Adjustments, based on the expected efficiency and effectiveness of the operation, may be considered by the designer.

6.4.2 Structural Elements

The vacuum assisted beds are proprietary devices and there are minor differences in the components offered by the various manufacturers. The listing below is a "generic" description of the common elements (15).

- A support structure, either a level concrete slab or level graded stone overlaying a sloped concrete slab upon which the media plates are placed.
- A concrete wall surrounding three sides of the bed and a bed closure system on the fourth wall, to allow for the containment of conditioned sludge on the plates and subsequent removal of dewatered sludge cake.
- A filtrate collection/drainage system between the media plates and the underlying concrete slab.
- Media plates, sealed around the edges to adjacent plates and to the walls of the concrete containment structure.
- Polymer feed and mixing systems to introduce dilute polymer into the sludge feed stream, and to provide proper flocculation.
- Sludge distribution piping, usually located on the walls of the containment structure.
- An air-tight filtrate sump adjacent to the bed and connected to the filtrate collection/drainage system.
- Float operated filtrate pumps located in the sump to convey collected filtrate to the treatment plant headworks.
- A vacuum system connected to the filtrate sump to induce a partial vacuum between the underdrainage system and the layer of conditioned sludge on top of the media plates.
- A source of high pressure, 480 to 830 kPa (70 to 120 psi), clean (no particulates) wash water for cleaning the surface of the media plates after removal of the dewatered sludge cake.
- Drains for the collection of the media plate washwater. If the drains are inside the beds they must have caps or seals to insure no leakage of applied sludge.
- A conveyance system to pump this washwater back to the treatment plant. With proper valving the washwater can flow to the vacuum filtrate sump. A separate washwater sump is sometimes used, and on occasion direct gravity flow to the headworks is possible.
- A control panel to operate all mechanical and electrical components of the system. These are typically designed to allow either manual or automatic sequencing of the operational cycle.
- In most cases an enclosure covering all mechanical and electrical components in the system, including the filtrate pump, the vacuum system, the polymer system, and the control panel.
- If year-round operation in cold climates, or operation during rainfall is necessary, an enclosure for the whole facility may be required, with heat addition during freezing weather.

- A front-end loader to allow for mechanical removal of the sludge cake from the bed.

6.4.3 Performance Expectations

There are reliable operating records for over 20 installations in the United States. These systems have dewatered a variety of sludges, including:

- Waste Extended Aeration -- Thickened and Unthickened
- Aerobically Digested Activated -- Thickened and Unthickened
- Anaerobically Digested -- Thickened and Unthickened
- Lime Conditioned Primary and Waste Activated
- Imhoff Tank.

Table 6-3 summarizes data from a recent U.S. EPA Technology Assessment (16) which examined performance at 13 operating systems. The cycle time in all cases was 24 hours, including cake removal and plate washing. The median polymer (liquid emulsion type) dose was 9 kg/Mg (20 lb/ton) dry solids.

Table 6-3. Performance Observed at 13 Vacuum Assisted Drying Beds

Relative Loading	Solids Loading	Total Solids	
		Sludge Feed	Sludge Cake
	kg/m^2/cycle	percent	percent
Low[a]	3.18	0.8	9-12
Median[b]	9.18	1.5-3.0	14-18
High[c]	37.84	8-10	30-35

kg/m^2/cycle x 0.2048 = lb/ft^2/cycle.
[a] Unthickened oxidation ditch sludge.
[b] Typical of aerobically digested WAS, thickened by decanting.
[c] Lime conditioned mixture of primary and thickened WAS.

Vacuum assisted drying beds are a proven technology. Section 6.4.5 compares the costs for vacuum assisted beds to conventional sand drying beds. It must be recognized that the end product from the two concepts will not usually be the same. The sand bed will typically produce a sludge cake exceeding 30 percent solids while the vacuum bed is typically operated to produce a liftable (12 to 15 percent solids) cake. If regulatory agencies require a higher solids concentration for disposal, an additional supplemental drying area may be needed for the vacuum assisted systems.

Practically all of the problems experienced with early systems (prior to 1984) have been solved by the manufacturers. Ongoing research and development has focused on prolonging the service life of the media plates and reducing their cost. The major market to date has been treatment systems with flows less than 7,500 m^3/d (2 mgd), however, the size of facilities giving serious consideration to this concept is gradually increasing.

In many cases it is possible to add multiple layers of polymer treated sludge, with decantation of supernatant between each. The bed can also be filled with water to the top of the media plates prior to sludge application, as described in Section 6-5. This will increase the gravity drainage rate and, in this case, the vacuum is only used in the final stage. Combining these techniques can very significantly increase bed capacity.

6.4.4 Operation and Maintenance

The typical operational cycle includes the following activities:

- At the start of the cycle the vacuum pumps are off, the bed closure system is in place, the filtrate pumps are on automatic, sufficient polymer is mixed or otherwise available, and drains in the bed (if present) are covered and sealed. The filtrate valve may be open or closed depending on the mode of operation. The media surface may be dry, wet, or covered with a thin film of standing water (< 1.5 cm), depending on the particular system.

- Valves on the sludge feed line are opened, with the polymer feed pump also operational.

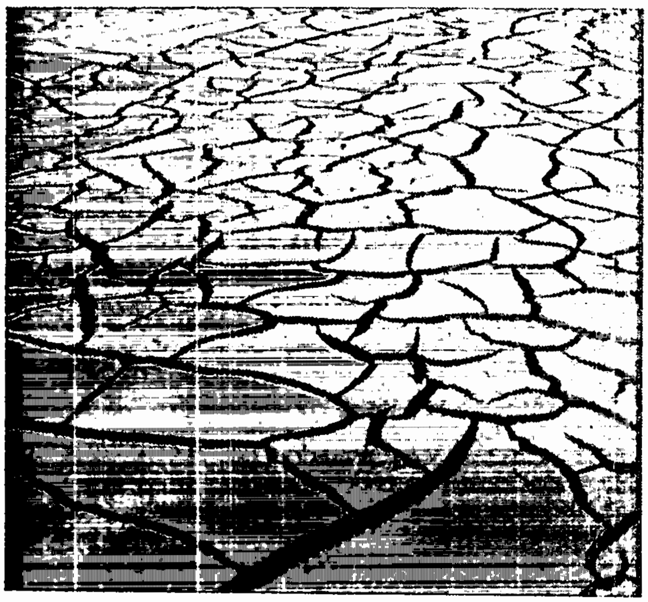
- If the cycle was started with the filtrate valve closed, it may be opened whenever the media plates become entirely covered with well flocculated sludge. Opening this valve allows gravity drainage to begin. If a good separation of sludge solids and supernatant occurs, it is also possible to decant the supernatant prior to opening the filtrate valve, as described in the previous section.

- When the desired volume of sludge has been applied to the bed, the polymer and sludge feed pumps are stopped.

- Gravity drainage is allowed to continue until the operator decides that the rate of filtrate collection is too slow. The time for this gravity drainage may range from 30 minutes to several hours after sludge application is completed.

- The operator starts the vacuum cycle at the end of gravity drainage. The vacuum sequence usually proceeds in discrete steps, beginning at 5 to 8 cm (2-3 in) Hg for about 1 hour, then increasing to 13 to 15 cm (5-6 in) Hg for another hour with a final step at 25 to 30 cm (10-12 in) Hg. This highest vacuum level normally continues until the sludge cake has dried sufficiently to crack and at this point the system vacuum level is lost. Figure 6-9 illustrates a "cracked" bed with the sludge ready for removal. Exact values and times vary from system to system.

Figure 6-9. "Cracked" bed, ready for sludge removal.



- An optional evaporative phase may be necessary to produce a liftable sludge cake. The minimum solids concentration to achieve this condition is about 10 to 12 percent for most sludges, but may be lower for certain others. The time required for evaporation is variable and can only be determined on a site specific basis.

- The bed closure system, if used, is removed to allow access for sludge removal. Typically a small tractor with a front loading bucket is used at all but the very smallest installations.

- The front-end loader cannot completely remove all of the sludge cake. The small amounts left on the bed must be removed manually with a shovel or a scoop. Diligent removal of this material is necessary to permit an optimum final cleaning of the media plates.

- The manual rinsing with hose and nozzle commences at the end of the bed furthest from the drains and progresses toward the drains. This media plate cleaning shares equally with the polymer conditioning as the most critical aspect of system operation. The plates must be scrupulously cleaned between each cycle of operation if progressive loss of plate permeability is to be avoided. This cleaning completes the operational cycle and the bed is ready for another charge of sludge.

Selection of polymer, the aging time, and the effectiveness of mixing and dosage control are variables subject to various degrees of operator control, and all of these factors strongly affect performance. The high cost of polymers makes overdosing a very expensive activity. In addition, overdosing may lead to progressive plate clogging and the need for special cleaning procedures to regain plate permeability.

Plate cleaning is critically important. If not performed regularly and properly, the media plates are certain to clog. The design should incorporate proper sizing and location of drains, sufficient water pressure, and selection of a satisfactory hose nozzle. The media plates will in time show some sign of decreased permeability, even with good maintenance, due to accumulation of oils and greases or other substances. Special cleaning measures are then required. Some of those used successfully include:

- High pressure, hot water cleaning

- Commercial grade hydrochloric acid at about 1 percent concentration

- Tri-sodium phosphate at about 0.25 to 0.5 percent

- Calcium or sodium hypochlorite at about 1 percent available chlorine

- Enzyme based cleaners.

6.4.5 Costs

Since vacuum assisted drying is a proprietary concept, the capital cost will vary with the manufacturer. The high rate operational cycle allows a compact system; land costs are not a significant factor. The manufacturer provides the media plates, pumps, polymer makeup/feed system, and system control panel at a typical cost of \$640-\$860/m^2 (\$60-\$80/ft^2) of bed surface area. The remaining capital costs are for concrete work, pipes and valves, electrical wiring, a control building for uncovered beds or a complete enclosure for the entire system if required, and a front-end loader for sludge cake removal. The range of total capital costs derived from a review of 29 systems are summarized in Table 6-4 (16).

Vacuum assisted drying beds (VADB) are normally compared to conventional sand drying beds. However, the sludge cake removed from the VADB is typically never as dry as that removed from sand beds and this difference must be recognized in any cost comparison. In some cases additional drying

Table 6-4. Capital Costs for Vacuum Assisted Drying Beds (1984 $)

Range	Capital Cost, $/m^2 Uncovered Beds	Covered Beds
Low	484	753
Median	1,485	1,614
Average	1,323	1,679
High	1,991	2,292

$/m^2 x 0.0929 = $/ft^2.

may be required prior to final disposal of the VADB sludge cake and that is not included in the comparisons in this section. As described in Section 6.2, the solids loading on sand beds is dependent on sludge characteristics, geographic location, and whether the beds are covered or not. Tables 6-5 and 6-6 (16) compare the relative capital and O & M costs for these two concepts.

Table 6-5. Cost Comparison of Vacuum Assisted Drying Beds vs. Sand Drying Beds

System Type	Average Capital Cost $/m^2
Vacuum Assisted Beds	
Uncovered	1,291
Covered (roof only)	1,475
Enclosed in a building	1,678
Sand Drying Beds	
Uncovered	121
Covered (roof only)	625
Enclosed in a building	1,000

$/m^2 x 0.0929 = $/sq ft.

Table 6-6. O & M Cost Comparison of Vacuum Assisted Drying Beds vs. Sand Drying Beds

Item	O & M Cost, $/Mg Sand Beds	Vacuum Beds
Labor	72	39
Polymer	-	29
Electricity	-	1
Front End Loader	1	1
Sand Replacement	7	-
Media Plate Cleaning	-	1
Media Plate Replacement	-	6
Total	80	77

$/Mg x 0.896 = $/ton.

The operational life of the systems compared in Table 6-5 is 20 years, with an assumed replacement of media plates after 10 years (possibly a very conservative assumption). The cost values in Table 6-5 do not include land costs which can be a significant factor for conventional sand beds. The land costs for possible additional drying of VADB sludge cake are also not shown.

The O & M costs in Table 6-6 are based on fuel and maintenance for the front-end loader, and assume an annual replacement of about 8 cm of sand on the sand beds and a 6-month chemical cleaning cycle for the vacuum bed media plates. The calculations assumed a 907 kg/d (2,000 lb/d) production of aerobically digested sludge, using 196 m^2 (2,112 ft^2) for the vacuum assisted bed and 1,859 m^2 (20,000 ft^2) for the sand bed.

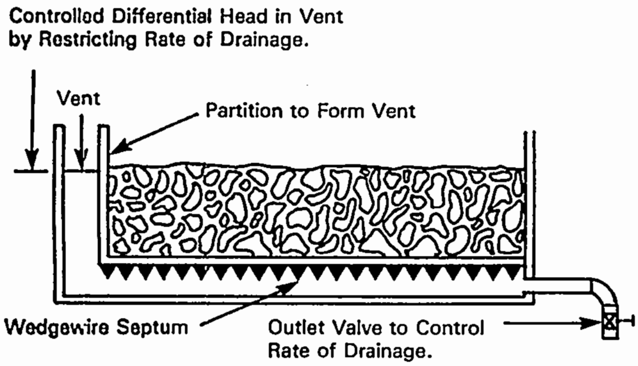
The cost effectiveness of these two technologies depends strongly on local climatic conditions, a critical factor in determining the loading rates on sand beds. The following relationships were derived from the information in Tables 6-5 and 6-6 and the related assumptions.

- An uncovered vacuum assisted bed will be more cost effective if the solids loading on an uncovered sand bed is less than 146 to 171 kg/m^2/yr (30-35 lb/ft^2/yr).
- A vacuum system with a roof will be more cost effective if the solids loading on a roofed sand bed is less than 317 to 342 kg/m^2/yr (65-70 lb/ft^2/yr).
- A vacuum system in a building will be more cost effective if the solids loading on a completely enclosed sand bed is limited to 440 to 464 kg/m^2/yr (90-95 lb/ft^2/yr).
- If the allowable solids loading on the sand bed exceeds the values given in the three categories above, then the conventional sand bed will be the more cost effective alternative.
- The solids content of the final sludge cake is also an important consideration. The vacuum assisted beds typically produce a liftable sludge (10-12 percent solids) while sand beds can achieve 50 percent solids or more. The hauling distance for final disposal and/or the necessary final solids concentration may require additional drying for the vacuum assisted product, thus increasing the total costs for this process.
- The use of polymers with sand beds, or the combined use of freeze thaw dewatering and polymer dewatering (as described in Section 6.3), should make uncovered sand beds competitive, even in colder climates.

6.5 Wedgewire Beds

The Wedgewire, or wedgewater, process is physically similar to the vacuum assisted systems described in the previous section. The media in this case consists of a septum with wedge shaped slots about 0.25 mm (0.01 in wide). This septum serves to support the sludge cake and allow drainage through the slots. Figure 6-10 illustrates the process.

Figure 6-10 Cross section of a wedge wire drying bed.



Initially, water enters the bed from beneath, and fills the bed to a depth of about 1 cm (0.4 in) above the media surface. Polymer conditioned sludge is then applied to the bed. During the initial phase the drain valve is closed so that the water and sludge stand on the bed. The valve is then partially opened to control the drainage rate for up to 2 hours. Following this controlled phase, the valve is opened fully and the sludge cake allowed to drain naturally. The initial static period with sludge on the flooded bed allows the sludge to settle to the media surface and form a filter zone. In addition, this establishes the potential for saturated flow conditions through the sludge and media. Drainage will proceed at a much higher rate under "saturated" conditions (devoid of air, so a small hydrostatic suction is exerted on the bed) compared to wet sludge resting on a dry surface.

6.5.1 Design Considerations

Since wedgewire systems are proprietary devices, loading criteria are developed in conjunction with the manufacturer. Bench scale or pilot units can be used to develop loading criteria and polymer dose for a particular sludge.

Typical sludge solids loadings range from 2 to 5 kg/m^2 (0.4 to 1 lb/ft^2) per operational cycle. The number of operational cycles per year will vary, depending on the type of system and other local conditions. On a routine, 24-hour operational cycle the annual loading could exceed 1,600 kg/m^2 (328 lb/ft^2), including allowances for maintenance downtime. This annual loading exceeds the loading on conventional sand beds by an order of magnitude. An enclosed and possibly heated facility would be needed to maintain such production in cold climates with extended periods of freezing weather. The wedgewire process appears best suited for smaller treatment systems, in locations with moderate climates, and where land area may be limited.

In many operational systems the sludge cake is removed soon after completion of the drainage phase to maintain high production rates. Typically, the sludge cake at this point will be 8 to 12 percent solids (after 24 hours) and is handlable, but still wet. Production of a drier sludge would require more time on the bed, or removal to a stockpile area for evaporative drying.

6.5.2 Structural Elements

The media was originally constructed of stainless steel but is now predominantly made of preformed polyurethane modules. The stainless steel requires additional support in the bed. The interlocking polyurethane modules are self-supporting and also create a shallow drainage plenum beneath the media surface. Either type of media can support small front-end loaders for removal of the sludge cake. Small systems have also used large tilting metal trays. In this case, when the sludge is ready for removal the whole bed is tilted to a steep angle and the sludge cake slides out.

Existing sand drying beds can be retrofitted for the wedgewire process, or a new concrete basin can be constructed.

6.5.3 Performance Expectations

According to manufacturers of wedgewire systems, polymer treated aerobically digested sludges can be dewatered to 8 to 12 percent solids within 24 hours and treated anaerobically digested sludges can be dewatered to 16 to 20 percent in the same time period. Polymer conditioning is necessary for most sludges and desirable for all. Without conditioning, the fines, particularly in aerobically digested sludges, may penetrate the media and either be lost with the filtrate or accumulate in the drainage plenum. To avoid solids build-up in the plenum, the floor of new concrete beds should be sloped to insure positive drainage.

6.5.4 Operation and Maintenance

The basic operation and maintenance requirements for wedgewire systems are similar to those described in Section 6.4 for vacuum assisted beds. Surface clogging is less likely with the wedgewire process but is still possible if routine cleaning is not performed properly.

Polymer conditioning is critical for successful performance. The polymer dosages required are similar to those used with the vacuum assisted beds described in Section 6-4. The typical sludge depth for a single application ranges from 10 to 25 cm (4-10 in). The optimum for a particular system will be determined with operational experience. In some cases, as also described in Section 6.4, it may be possible to apply multiple sequential layers with decantation of the supernatant prior to starting the drainage phase.

There have been reports of damage to the plastic surfaces when front-end loaders have been improperly used to remove the sludge cake. The proper procedure requires driving straight in and backing straight out. Sharp, skidding turns can cause structural damage to the molded polyurethane surfaces.

It is important for the operator to carefully manage the initial controlled drainage rate to insure maximum water flow during this phase. If the rate is too slow the total cycle time will have to be increased, and if the rate is too high, complete drainage may not occur. The manufacturer's drainage recommendations can be used initially and then modified as necessary with operational experience.

6.5.5 Costs

Land is not usually a significant factor in the capital costs of wedgewire systems, unless additional land area is needed for further drying of the sludge cake. The other construction costs will depend on whether existing sand beds can be retrofitted instead of constructing entirely new basins. Construction costs in the latter case might range from $1,000 to $1,900/m^2 ($93 to $177/ft^2). Operating costs may be slightly less than the vacuum assisted systems described in Section 6.4, but will still fall in the same range. The comparisons in Section 6.4.5 to sand beds should also be approximately applicable to wedgewire systems.

6.6 Sludge Lagoons

A distinction must be made between sludge drying lagoons and sludge lagoons primarily intended for storage. Some drying occurs in storage lagoons but the primary intent is to provide temporary or semi-permanent storage.

Drying lagoons are operated on a regular cycle to dewater sludges. A typical operational cycle includes the following activities:

- Well stabilized liquid sludge is pumped into the lagoon, over a period of several months or more.

- Supernatant is decanted, either continuously or intermittently, from the lagoon surface and returned to the treatment plant.

- Filling and decanting operations are continued until the design depth of sludge is reached.

- The surface crust is repeatedly broken up and/or removed during the drying period.

- Dewatered sludge is removed with some type of mechanical removal equipment.

- Maintenance and repair is performed while the lagoon is empty and then the filling cycle is repeated.

The complete cycle for a single lagoon typically takes from less than 1 to 3 years, depending on the final solids concentration required, local climate, the depth of sludge applied, and management practices (17). All sludge should be stabilized prior to addition to the lagoon to minimize odor problems. Occasional odors, flies and mosquitos may still be a problem, so a remote site is essential.

6.6.1 Design Considerations

Until recently, sludge lagoons were often located in soils with at least moderate permeability to take advantage of subsurface drainage and percolation. That practice is now the exception rather than the rule in most of the United States due to more stringent environmental and groundwater protection regulations. If a groundwater aquifer with drinking water potential exists beneath the site, it may be necessary to line the lagoon or otherwise restrict significant percolation. Unless a sand bottom and underdrains are then installed, the only sludge dewatering mechanisms left are decanting supernatant and evaporation.

In effect, the sludge drying lagoon is similar in concept to a deep sand drying bed with restricted drainage. The depth of sludge in the lagoon might be 0.7 to 1.4 m (24 to 48 in) as compared to 0.3 m (12 in) for the sand bed. The recommended solids loading for the drying lagoons is 36 to 39 kg/yr/m^3 of lagoon capacity (2.2 to 2.4 lb/ft^3/yr). A minimum of two cells is essential, even at very small systems, to insure availability of storage space during cleaning, maintenance or emergency conditions.

Evaporation and decantation are usually the dominant pathways for water even if an underdrainage network exists. The required lagoon surface area depends on the temperature, precipitation, and evaporation rates for the local area. Equations 6-8 to 6-12 in Section 6.7 can be used to estimate surface area requirements, or assuming that standing water is routinely decanted, the design calculations for evaporation are similar to Equations 6-1 to 6-4. The evaporation procedures in Reference 18 to complete retention ponds can also be used. The water to be removed from the sludge lagoons is the required portion of the sludge moisture content plus that portion of precipitation that will infiltrate the sludge mass rather than be removed as supernatant.

The dependence on evaporation tends to favor arid and semi-arid climates for this dewatering process. However, the Metropolitan Sanitary District of Greater Chicago, the Milwaukee Metro Sewerage Authority, and the City of Philadelphia have all successfully operated large scale sludge drying lagoons in cool humid climates (19).

It is possible to facilitate drying with a device that consists of a tractor with a helical screw in front to push sludge aside and mix it. This helps to open up the dried top layer and expose the wet material below.

6.6.2 Structural Elements

The retaining walls for drying lagoons are typically earthen dikes 0.7 to 1.4 m (2 to 4 ft) high with a side slope of 1:3. The lagoon is typically rectangular in shape to facilitate sludge removal. Required equipment includes: sludge feed lines and pumps, supernatant decant lines, and sludge removal equipment. The last can include trucks, front-end loaders, bulldozers, or draglines, depending on the size of the operation.

6.6.3 Performance Expectations

Solids concentrations in the range of 15 to 40 percent are expected in the sludge removed from the lagoon; concentrations can be higher in arid climates. These lagoons share a common problem with other air drying processes in that a surface crust forms early in the evaporative stage, which then restricts further evaporative water losses. This problem is minimized with the paved drying beds described in Section 6.8 that use mechanical equipment to move around the bed to turn and mix the sludge. Similar equipment and procedures can be used in drying lagoons if the depth of sludge permits. Floating devices can also be used. Larger scale facilities have used a cable and scraper system as shown in Figure 6-11.

6.6.4 Operation and Maintenance

The routine operational activities consist of sequential sludge applications and decantations until the lagoon contains the design volume of sludge. The periodic break-up or removal of the surface crust then insures continued evaporation. Sludge removal is labor intensive but occurs infrequently. Maintenance activities include care of equipment and dikes and control of dike vegetation. Some sludge drying lagoons may require insect and odor control. The labor requirements for sludge drying lagoons are shown in Figure 6-12.

6.6.5 Costs

The capital cost for drying lagoons is significantly influenced by the cost of land at the project site. Other major factors include construction of the dikes, sealing the bottom (if required), underdrainage (if used), and the other structural elements described in Section 6.6.2. The construction costs for the lagoon (with earthen dikes) are similar to the costs for sludge storage lagoons, or wastewater treatment ponds. Appendix A-32 in Reference 8 can be used to estimate these costs. The other capital costs depend on the intended methods for sludge loading and removal, and should be determined on a case-by-case basis. The major O & M costs are for labor, fuel, and maintenance of sludge removal equipment. Figure 6-12 (20) can be used with prevailing wage rates to estimate labor costs. The remaining O & M costs will depend on the equipment and procedures used and must also be determined on a case-by-case basis.

6.7 Paved Beds

Until recently, paved beds used an asphalt or concrete pavement on top of a porous gravel subbase. Unpaved areas, constructed as sand drains, were placed around the perimeter or along the center of the bed to collect and convey drainage water. The main advantage of this approach was the ability to use relatively heavy equipment for sludge removal. Experience showed that the pavement inhibited drainage, so the total bed area had to be greater than that of conventional sand beds to achieve the same results in the same time period.

Recent improvements to the paved bed process utilize a tractor-mounted horizontal auger, or other device, to regularly mix and aerate the sludge (21). This mixing and aeration breaks up the surface crust that inhibits evaporation, allowing more rapid dewatering than conventional sand beds. Some of the equipment was originally developed for composting operations but serves equally well for paved bed dewatering. Underdrained beds are still used in some locations, but the most cost effective approach in suitable climates is to construct a low cost impermeable paved bed and depend on decantation of supernatant and auger/aeration mixing for evaporation to reach the necessary dewatering level. Figure 6-13 shows a bed of this type.

6.7.1 Design Considerations

The critical design parameter for paved beds, as with sand beds and drying lagoons, is the surface area required to dewater the sludge to the specified solids level in the specified time. Since drainage is not a factor in many modern paved bed designs, the only ways water can be removed is through decantation and evaporation. These water losses will depend on the same factors described in Section 6.2, but with paved beds the use of the mechanical auger/aerator sustains evaporation near the maximum potential for sludge. Paved beds can be used in any location, but since evaporation provides the major pathway for water loss, they work best in warm, arid and semi-

Figure 6-11 Cable and scraper system for sludge drying lagoons.

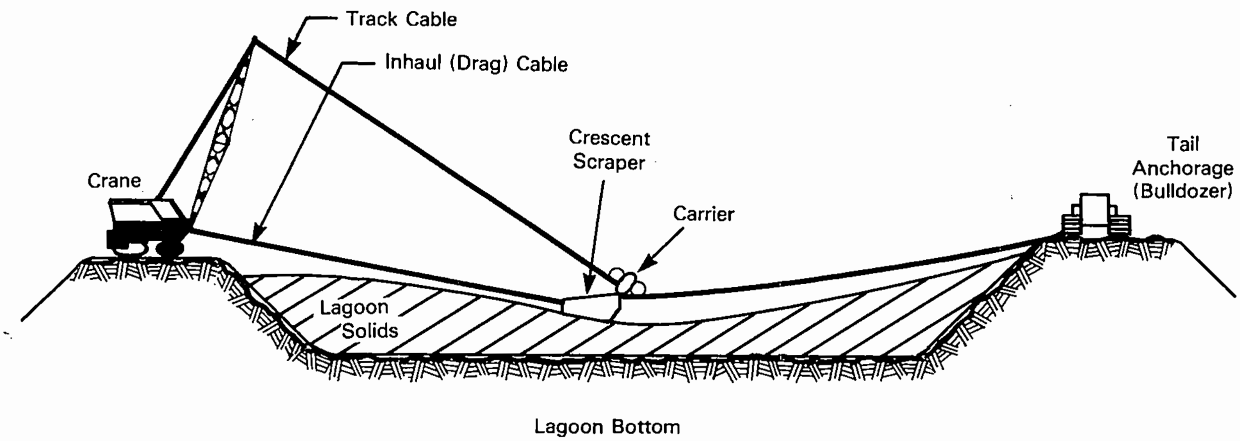
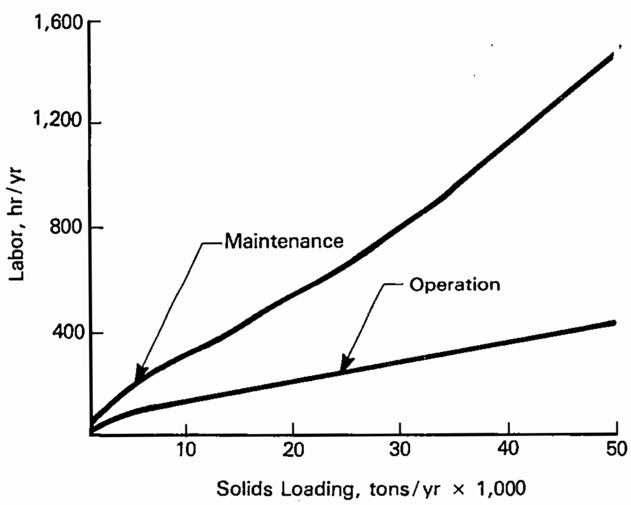


Figure 6-12. Labor requirements for sludge drying lagoons.



arid climates. Assuming the same degree of effort with the auger/ aerator, the design solids loading on a bed, or the bed area will be directly related to the potential evaporation, and precipitation in the local area. The design loading rate for the system in Roswell, NM, is 244 kg/m^2/yr (50 lb/ft^2/yr); the loading during a pilot test in Wichita, KS, was 127 kg/m^2/yr (26 lb/ft^2/yr). In more humid climates the allowable loading might be even lower (3).

As shown in Figure 6-13, these completely paved beds incorporate devices to draw off the supernatant, and with some sludges it may be possible to draw off 20 to 30 percent of the water in this manner. If the sludge has particularly good settling characteristics, it may be possible to use several fill and decant cycles prior to the evaporative stage. The rate of evaporation for a particular site can be determined with small scale pilot studies or assumed as a fraction of the pan evaporation rate for water in the local area (usually routinely available). A study in New Mexico (3) indicated that the evaporation rate from mixed and aerated sludge was about 58.7 percent of the free water pan evaporation for the site. That relationship should be generally valid for other locations also. At large scale projects, where land costs can be very significant, a pilot test to determine this ratio should be used to optimize the design.

The water losses and bed area required for a paved bed system can be determined with Equations 6-8 to 6-12.

$$W_O = (1.04)(S)\left[\frac{1 - s_0}{s_0}\right] \qquad (6-8)$$

where,

W_O = total water content in applied sludge, kg/yr (lb/yr)

1.04 = assumed specific gravity of sludge solids

S = annual sludge production, dry solids, kg (lb)

s_O = dry solids in applied sludge, percent as decimal

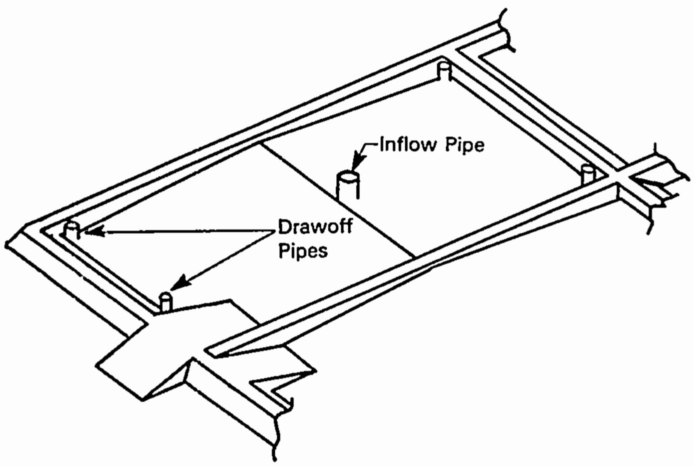
The water content after decantation is given by:

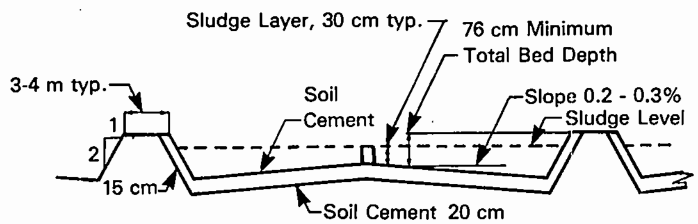
$$W_D = (1.04)(S)\left[\frac{1 - s_d}{s_d}\right] \qquad (6-9)$$

where,

W_D = total water remaining after decantation, kg/yr (lb/yr)

Figure 6-13 Paved sludge drying bed designed for decantation and evaporation.





Cross Section View

s_d = dry solids in sludge after decantation, percent as decimal

The water content to be removed by evaporation is given by:

$$W_E = W_D - (1.04)(S)\left[\frac{1 - s_e}{s_e}\right] + (P)(A)(1{,}000) \qquad (6\text{-}10)$$

where,

W_E = water to be evaporated after decantation, kg/yr (lb/yr)

s_e = dry solids required after evaporation, percent as decimal

P = annual precipitation, m (ft)

A = bed area, m^2 (sq ft)

In US units, the final term becomes: (P)(A)(62.4). If the system does not allow decantation, use W_O in equation 6-10 instead of W_D.

The evaporation rate for a given location is given by:

$$R_e = (10)(k_e)(E_p) \qquad (6-11)$$

where,

R_e = evaporation potential for sludge on a mixed and aerated paved bed, kg/m^2/yr (lb/sq ft/yr)

k_e = reduction factor for sludge evaporation vs. a free water surface, percent as decimal
= 0.6 (pilot test to determine this value is recommended for large projects)

E_p = free water pan evaporation rate, cm/yr (ft/yr)

In US units, the 10 becomes 62.4.

The area required for a paved drying bed system can be estimated by combining Equations 6-10 and 6-11.

$$A = \frac{(1.04)(S)[(1-s_d)/s_d - (1-s_e)/s_e] + (P)(A)(1000)}{R_e} \qquad (6\text{-}12)$$

where,

A = paved bed area, m^2 (sq ft, use 62.4 instead of 1,000)

The example below demonstrates the application of this procedure.

Assume:

S = 365,000 kg/yr total dry solids in sludge produced

s_d = 15 percent

s_e = 35%

P = 0.5 m/yr

E_p = 127 cm

Use equation 6-11 to determine evaporation rate:

R_e = (10)(0.587)(127) = 745 kg/m^2/yr

Use equation 6-12 to determine the total area required:

A = [(1.04) (365,000) (5.67 - 1.86) + (0.5) (A) (1,000)] ÷ 745.5

A = 1,940 + 0.67 (A)

A (1 - 0.67) = 1,940

A = 5,891 m^2 (0.5891 ha, 1.46 ac)

The solids loading for this case would be 62 kg/m^2yr.

The total design area should be divided into at least three beds for all but the smallest operation to provide operational flexibility. A detailed month-by-month analysis of weather records and expected sludge production rates will determine the optimum number of beds required. It may not be necessary to use all of the beds in the hot dry months. For example, the system designed for Roswell, NM has a total of seven beds, six of which need to be used in December; only three are required in June due to increased evaporation and decreased sludge production. To insure a conservative design, the equations and the example presented above assume that all precipitation that falls on the bed must be removed by evaporation. For small systems, it may be advantageous to plan the orientation of the bed for reception of maximum solar radiation.

6.7.2 Structural Elements

Paved beds have been constructed with concrete and asphalt pavement, with and without drains. However, the most economical approach may be to use soil cement as the paved surface as shown in Figure 6-13. Other structural features are also shown in the same figure. Information on construction of soil cement pavements can be obtained from the Portland Cement Association.

A long rectangular configuraton improves efficiency by reducing the time required for turning the auger/aerator vehicle. A variety of inlet and decantation structures are also possible. The minimum total depth of the bed is about 0.8 m (2.6 ft) to provide some freeboard above the typical 30-cm (12-in) sludge layer. In some systems up to 1 m (3 ft) of liquid sludge is applied in the initial layer and the freeboard must be correspondingly increased.

Other major system components include the sludge and decantation piping, and the auger/aerator vehicles. A variety of vehicle sizes and configurations are available and the designer should seek the assistance of the manufacturers in determining the optimum sizes and number for a particular operation.

6.7.3 Performance Expectations

The use of digested, or otherwise stabilized, sludge is necessary to avoid odor complaints and to satisfy regulatory requirements for final sludge disposal. The decantation phase might require 2 to 3 days for sludge settling and 1 to 2 days to decant each increment of sludge added, depending on the sludge characteristics. If drainage is allowed by the design, it should also be essentially complete during the time allowed for sludge settling and decantation.

The final evaporative drying period will depend on the climatic conditions occurring after the sludge is applied and on the regular use of the auger/aerator equipment. Solids in the range of 40 to 50 percent can be achieved in 30 to 40 days in an arid climate, for a 30-cm (12-in) sludge layer, depending on the time of year and the effectiveness of decantation (3). A 1-m (3-ft) sludge layer in the same climate might require 100 to 250 days to reach 50 percent solids, depending on when the sludge was applied.

6.7.4 Operation and Maintenance

The major operational tasks are sludge application, decantation, mixing and aeration, and sludge removal. Depending on the size of the operation and the time of year, the sludge on the bed should be mixed several times a week to maintain optimum evaporation conditions. Labor requirements at the

Roswell, NM system are estimated to be about 0.3 hr/yr per Mg of dry solids processed (3). Maintenance requirements include routine care of the auger/aeration equipment, the sludge pumping and piping network, the decantation piping, and the bed and dikes. If the site experiences freezing weather in the winter months, the valves and pumps in the system need to be protected and checked periodically during the critical freezing periods.

6.7.5 Costs

Capital costs are strongly influenced by the cost of land at the project site. Other major capital costs include the containing walls and pavement, application and decantation piping (and drainage piping if used), the auger/aerator, and the sludge removal equipment. In many cases, the same vehicle can be used for both tasks. The major O & M costs are labor and fuel for the equipment. Table 6-7 (3) compares the cost of a paved bed operation to conventional sand beds in the same location.

Table 6-7. Estimated Cost Comparison of Paved Beds vs. Conventional Sand Drying Beds

Item	Sand Drying Bed	Paved Bed w/Auger/Aerator
Number of Beds	16	7
Total Area, m^2	60,600	26,200
Solids Loading, $kg/m^2/yr$	108	243
Labor, hr/yr	8,580	1,700
Capital Costs, $	1,465,000	520,000
O & M Costs, $/yr	100,000	25,000
Total Present Worth, $	2,500,000	780,000

6.8 Other Innovative Processes

A number of other processes, which do not fit directly into the categories previously discussed, have been proposed and tested at both the pilot and full-scale level. In some cases the distinctions are minor. The "Solar" sludge drying beds used in the arid southwestern United States differ little in concept from drying lagoons discussed in Section 6.6 or the paved beds described in Section 6.7. In all cases the key to success is the mixing and turning of the sludge and break-up of the surface crust. The evaporative losses will be very rapid in arid climates. There are two other processes that are unique; one incorporates additives of various types and the other utilizes an underdrained sand bed and growing reeds or bulrushes to dewater the sludge.

6.8.1 Additives

In some cases additives such as sawdust, wood chips, etc., are mixed with the sludge as a preparatory step for composting or vermistabilization. These processes are usually described as stabilization processes but a significant degree of dewatering does occur with both.

In a composting process, the initial sludge solids concentration might be about 20 percent. Wood chips, sawdust or some other bulking agent is added so the solids content of the mixture is between about 40 and 45 percent. The solids content of the compost product following composting might approach 50 to 60 percent (22). There is considerable moisture loss to the atmpshere. Most of the energy for evaporating water comes from the aerobic oxidation of the sludge, but contact with the air by forced aeration or turning of windrows is needed to carry the moisture away.

Sawdust is also used as a bulking agent in a process that uses earthworms for sludge stabilization and dewatering (vermistabilization). In this case, thickened sludge (3-4 percent solids) is sprayed onto beds containing sawdust and earthworms. Typical sludge loading rates are 85-90 $kg/m^2/yr$ (17-18 $lb/ft^2/yr$), which is near the low end of the range for conventional sand beds.

After 6 to 12 months the mixture of earthworms, castings, and sawdust is removed and the earthworms separated (by screening) for use in the next cycle. The solids concentration of the stabilized material ranges from 15 to 25 percent. This process has been demonstrated at the pilot scale level at Lufkin, TX, and elsewhere (23). A heated enclosure over the beds would be required in all but the warmest climates to sustain activity during the winter months. Laboratory scale experiments at Cornell University successfully stabilized and dewatered raw sludges with this process (24). A large scale version of this process might be a cost-effective alternative since thickening, digestion, conditioning, and dewatering are all eliminated.

In the Pacific Northwest, sawdust has also been used as a bulking agent for sludge drying in a process similar to a paved bed operation. The use of sawdust or other agents will only be economical when the materials are locally available at little or no cost.

The reed bed process combines the elements of an underdrained sand bed with a dense stand of vegetation to obtain sludge dewatering. Most of the operational reed beds have been planted with the common reed Phragmites but rushes or cattails could also be used.

The bed is actually constructed as a deep trench and lined to prevent exfiltration. A 25-cm (10-in) layer of washed gravel encloses the underdrain pipe, and is overlain by a 10-cm (4-in) layer of sand. The root stock of the reeds is planted on 30-cm (12-in) centers on the gravel layer, at a depth of about 10 to 15 cm (4-6 in). The bed is flooded with water to a depth of about 10 cm (4 in) for several weeks to encourage plant development. The freeboard above the sand layer is at least 1 meter to provide for long-term sludge storage. Sludge is not applied until the plants are well established.

The vegetation plays an essential role in the dewatering process. The root system absorbs water which is then lost to the atmosphere via evapotranspiration. More importantly, the penetration of the plant stems and the root system maintains a permanent pathway for continuous drainage of water from the sludge layer. Reeds and similar plants have the capacity to transmit oxygen from the leaves to the roots so there are aerobic microsites (adjacent to the roots) in an otherwise anaerobic environment that assist in stabilization and mineralization of the sludge.

Stabilized, thickened sludge, at about 3 to 4 percent solids is applied in 10-cm (4-in) layers. Solids concentrations less than 3 percent are not cost effective for optimum bed use and concentrations above 4 percent will not flow properly and will not be uniformly distributed within the dense vegetation. A layer can be applied about every 20 days with warm, dry weather conditions.

An operating system in Washington Township, NJ (25) was designed for an annual loading of 3.5 m (11.5 ft) of aerobically digested sludge at 3 percent solids, solids loading of 100 $kg/m^2/yr$ (20 lb/sq ft/yr). The average loading on 16 operational systems in New Jersey, New York, and North Carolina is about 81 $kg/m^2/yr$ (17 lb/sq ft/yr) which is at the low end of the range for conventional sand beds. The final annual layer of dried and further stabilized sludge (at about 90 percent solids) will be about 10 cm (4 in) deep and this residue can be left in place. A 10-year operational cycle has been planned for several systems in New Jersey (26). At the end of this period the accumulated sludge and the sand layer are removed. A new layer of sand is installed and new vegetation planted if necessary. An annual harvest of the vegetation is recommended, when the plant is dormant but before the leaves have been shed. The harvested material can be burned, composted, or otherwise disposed of.

These systems have been successfully operated in New Jersey on a year-round basis, with only 20 to 30 days downtime for adverse weather conditions. Since the dewatering benefits will be minimal during the dormant season for the plants and during prolonged freezing weather, it is likely that a longer downtime will be required for locations with more severe winters than New Jersey.

Multiple beds are required for every installation. With a 10-year cycle, a minimum of 12 beds would be necessary to allow for one out of service each year and one for emergencies. When a bed is to be cleaned, sludge applications are stopped for that bed in early spring, the vegetation is harvested in early fall, and the sludge residue and sand are removed by early winter. The dried sludge removed from the bed is similar in character to composted sludge with respect to pathogen content and stabilization of organics, due to the long detention times combined with the final 6-month rest period.

The major advantage of the reed bed process is the infrequent need for sludge removal and bed cleaning. The major disadvantage is the need for the annual vegetation harvest. However, the total volume of harvested vegetation and sludge residue on a 10-year operational cycle is still less than the sludge cake volume requiring disposal if the same amount of sludge were dried on a conventional sand bed.

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When an NTIS number is cited in a reference, that reference is available from:

National Technical Information Service
5825 Port Royal Road
Springfield, VA 22161
(703) 487-4650

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Chapter 7
Mechanical Dewatering Processes

7.1 Introduction

Some conditions favoring the use of mechanical dewatering are as follows:

- *Aesthetics*: Developed areas, where land is a premium and the use of open air drying might be offensive, are prime candidates for the use of mechanical dewatering.
- *Climate*: Adverse weather conditions are not conducive to non-mechanical drying methods.
- *Costs*: Hauling liquid material a significant distance is not as cost-effective as hauling dewatered material.
- *Site limitations*: Lack of available land within an economical liquid haulage distance (or sludge that is not suitable for land applications) will favor the use of mechanical dewatering.

While vacuum filtration remained the most popular method of sludge dewatering into the early 1970s, this practice was sharply curtailed by the end of the decade. No significant improvements in vacuum filtration have occured recently, and none are currently foreseen. Improvements in belt presses and centrifuges resulted in a shift to these devices; they were more cost-effective and generally produced better results than vacuum filtration. The oil shortage in the mid-seventies produced a renewed interest in filter presses, especially in combustion operations where fuel costs had increased four- to eight-fold in a matter of two to three years. The fact that both belt presses and centrifuges were more efficient in the use of polymers weighed heavily in their favor. Recently, filter press operations have had success with polymers. This development should lead to further interest in filter presses, particularly for composting, combustion, and restricted landfills (moisture limited).

Lime and ferric chloride conditioning have become a less favored option than polymer conditioning for several reasons. With these conditioning chemicals, operating and maintenance (O&M) costs are higher and the operation is less clean and more labor intensive. The seventies saw an increase in the cost of lime, as much as 200-300 percent. Further, lime and ferric chloride conditioning increases the gross solids for disposal by 20-25 percent, whereas polymer produces a negligible increase in solids quantity.

The comparative results obtained with the various methods of mechanical dewatering can vary from sludge to sludge. It is realistic to compare data of different devices only when they have been tested on the same sludge, during the same time period, and at the same relative level of capacity. That is, one should not compare results of one mechanical dewatering unit operating at 50 percent of its capacity with one at 100 percent of its capacity.

Still, it is possible to compare mechanical dewatering devices operating at maximum practical capacity in relative terms to each other in terms of cake solids, chemical costs, and solids recovery. This comparison is shown in Table 7-1. There are not detectable factors without testing that can result in substantial differences in cake and polymer costs between mechanical dewatering devices. For example, a belt press may produce a range of cake solids between 19 and 30 percent TS on a mixture of 50 PS:50 WAS when a large volume of data on sludges from several sources is compared. Because of the great variability between sludges, larger projects and those requiring a drier cake should be field tested whenever possible. The implication is not that it is all right to blunder on small projects. Rather, the cost of testing is a high fraction of the total capital cost for small projects and correction of errors is frequently relatively easy.

7.2 Belt Filter Presses

7.2.1 Introduction

Belt filter presses (BFP) have been used in Europe since the 1960s and in the United States since the early 1970s. They were initially designed to dewater paper pulp and were then modified to dewater sewage sludge. The European models were the first used in the United States. However, the difference between U.S. sludge and European sludge led to performance problems (low cake solids and poor solids capture). American manufacturers began

Table 7-1. Comparative Mechanical Dewatering Performance

Dewatering Unit	Cake Solids[1]	Recovery	Polymer Cost[3]
	% TSS	% TSS	%/ton DS
Belt press	X	90-95[2]	Y
Centrifuge	X ± 2	90-95[2]	0.8 Y
Vacuum filter	X - 4	85-90	0.9 Y
Filter press - Lo P	X + 8	98+	1.1 Y
Filter press - Hi P	X + 10	98+	1.1 Y
Filter press - Dia P	X + 12	98+	1.1 Y
Screw press	X - 2	90+	1.2 Y
Low press drum/belts	X - 10+	90+	0.8 Y

[1] Relative to belt press, X denotes base level.
[2] Controlled by polymer dosage.
[3] Relative to belt press, Y denotes base level.

building and selling presses in the United States. However, the early American-made models were also plagued with many problems, mainly mechanical failures of rollers and bearings. Since the American units were designed with the same principles used for the design of belt conveyors, they were much lighter than their European counterparts. The bearings and roller shafts were unable to withstand the forces generated during the dewatering operation.

By the late 1970s, American manufacturers made significant improvements in their units. For example, they followed the European practice of using bearings rated at an L_{10} life of 100,000 hours. Further, they increased the diameter of the roller shafts and improved the materials of construction, thus reducing the number of failures considerably. Since these improvements have been made, the use of belt filter presses has increased. With the newest models on the market, mechanical failures are limited and process performance has improved. Although there are now about 20 belt filter press suppliers in the United States, only about 5 are considered to be major manufacturers.

7.2.2 General Description

Belt filter presses are designed on the basis of a very simple concept. Sludge sandwiched between two tensioned porous belts is passed over and under various diameter rollers. For a given belt tension, as roller diameter decreases, an increased pressure is exerted on the sludge, thus squeezing out water. Although many different designs of belt filter presses are available, they all incorporate the following basic features:

- Polymer conditioning zone
- Gravity drainage zone
- Low pressure zone
- High pressure zones.

Figure 7-1(1) shows these zones identified on a simplified schematic of a belt filter press.

The polymer conditoning zone can be a small tank, approximately 265-379 l (70-100 gal) located 0.6-1.8 m (2-6 ft) from the press; a rotating drum attached to the top of the press; or an in-line injector. Each press manufacturer usually supplies the polymer conditioning unit with the belt filter press.

The gravity drainage zone is a flat or slightly inclined belt which is unique to each press model. In this section sludge is dewatered by the gravity drainage of the free water (interstitial water in the sludge slurry). The engineer should expect a 5 to 10 percent increase in solids concentration in the gravity drainage zone from the original feed sludge (2,3). For example, a primary/waste activated sludge mixture fed to the press at about 3.0 percent solids will be about 10-12 percent solids at the end of the gravity zone. Problems such as sludge squeezing out from between the belts and blinding of the belt mesh can occur if the sludge does not drain well in this zone. This free water drainage is a function of sludge type, quality, conditioning, screen mesh, and the design of the drainage zone.

The low pressure zone, also called the wedge zone by some manufacturers is the area where the upper and lower belts come together with the sludge in between, thus forming the sludge "sandwich." The low pressure zone is very important since it prepares the sludge by forming a firm sludge cake which is able to withstand the shear forces within the high pressure zone.

In the high pressure zone, forces are exerted on the sludge by the movement of the upper and lower belts, relative to each other, as they go over and under a series of rollers with decreasing diameters. Some manufacturers have an independent high pressure zone which uses belts or hydraulic cylinders to further increase the pressure on the sludge (see Fig. 7-2), thus producing a drier cake. A dry cake is especially important for plants that use incineration as the final disposal method and need the driest cake possible.

7.2.3 Theory of Operation

The high pressure zone is critical to good press performance (high cake solids and recoveries). A design manual prepared by Rubel and Hager, Inc. (4) contains models to describe the various effects in a typical high pressure zone (see Figure 7-3). This design manual provides these equations to familiarize the engineer with the belt filter press design process. The equations can be used to calculate the following parameters:

- Pressure on the sludge cake due to drive torque (force required to pull the belt through the press):

Figure 7-1. Simplified schematic of a belt filter press.

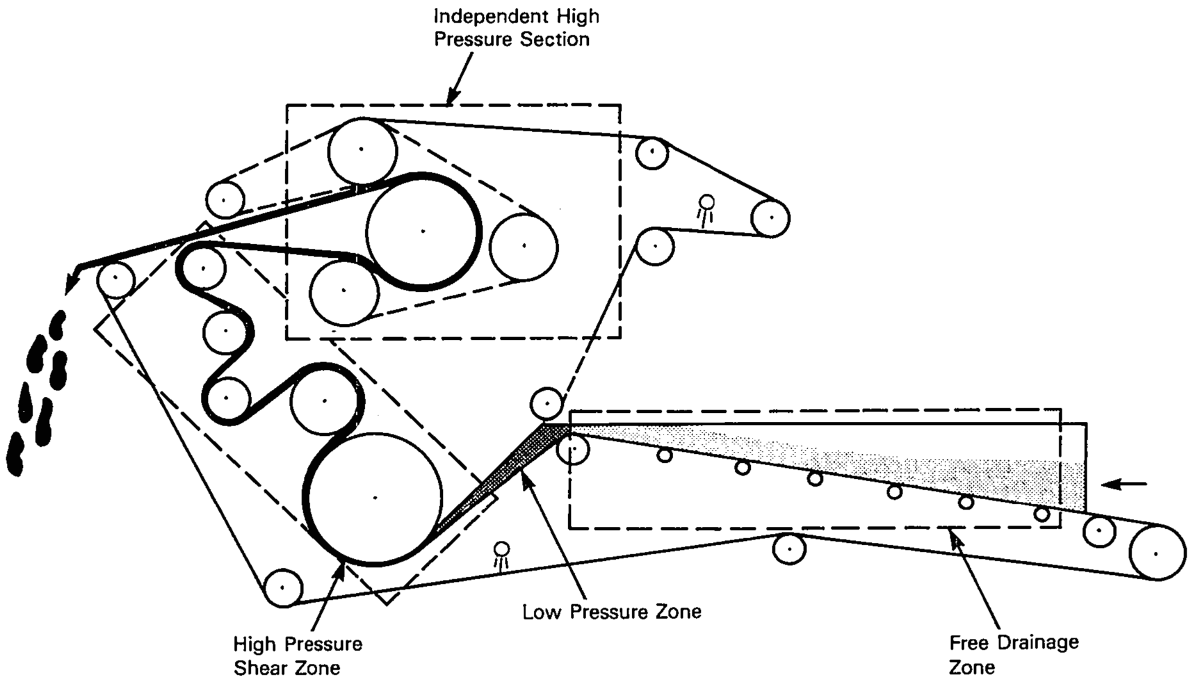
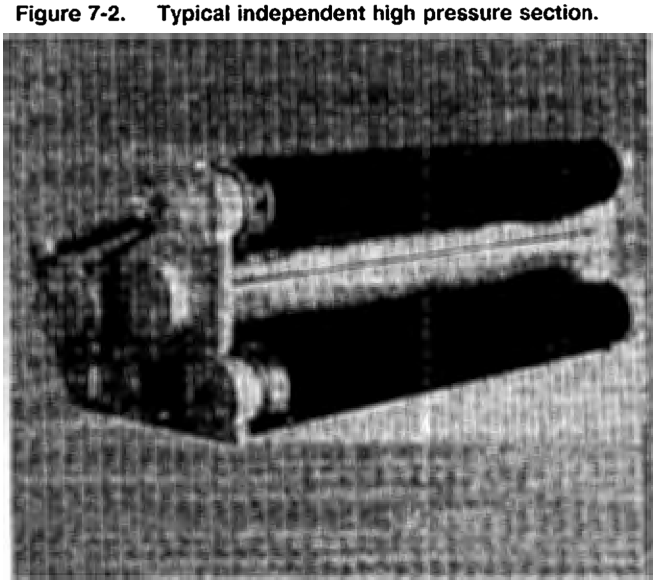


Figure 7-2. Typical independent high pressure section.



$$psi_1 = 2F_1/D = 1{,}700\ HP'/(D)\ (fpm) \qquad (7\text{-}1)$$

where,

psi_1 = maximum pressure on the sludge cake due to F_1
F_1 = lb of force due to drive torque per inch of belt width
D = roller diameter, in
HP' = drive horsepower per inch of belt width per belt
fpm = belt speed, ft/min

- Pressure on the sludge cake due to belt tensioning (for presses that use pneumatic or hydraulic cylinders to tension the belts)

$$psi_2 = 2F_2/D + 2P \cos [a/(DWY/2)] \qquad (7\text{-}2)$$

where,

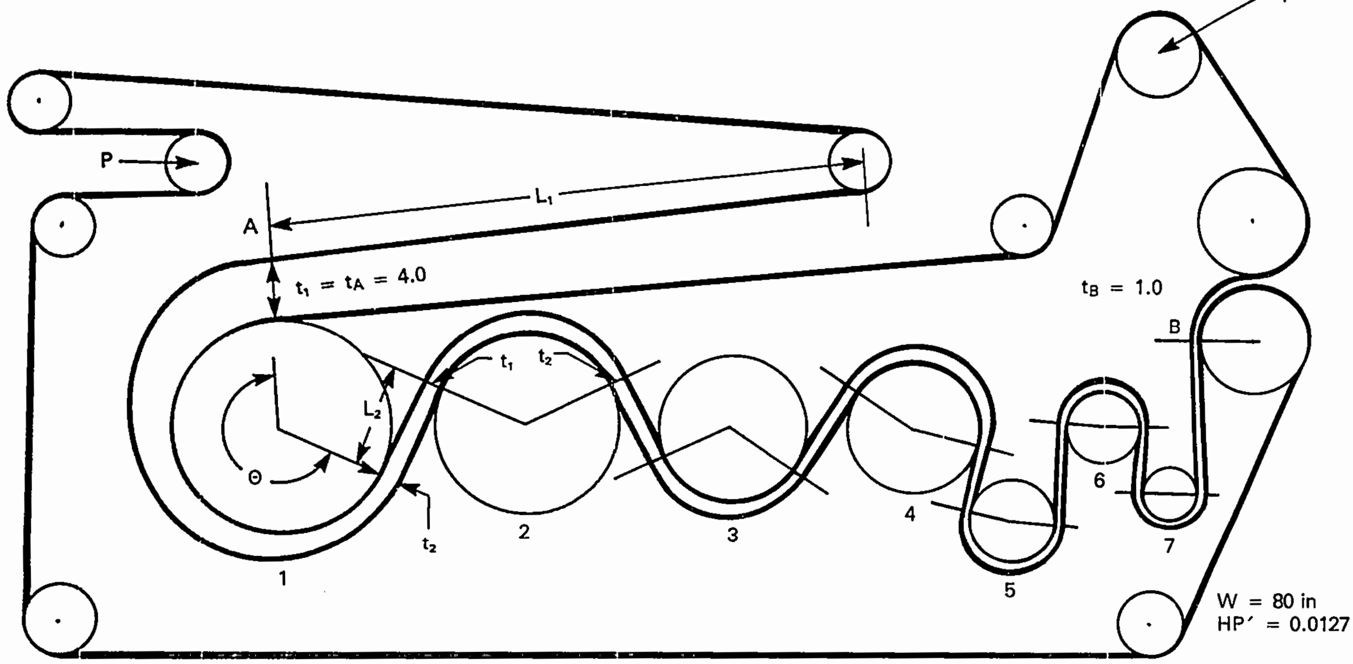
psi_2 = average pressure on sludge cake due to F_2
F_2 = lb of force due to take-up tension per inch of belt width - required to prevent slack belts and to provide traction for the drive rolls
P = resultant force from tensioning roller actuator. It is the pressure (force) you set and can easily measure.
a = angle between belt force resultant and actuating cylinder axis
D = diameter of roller, in
W = active belt width, in
Y = belt wrap angle at take-up roller

- Pressure on the sludge cake due to belt elasticity

$$psi_3 = 2F_3/D \qquad (7\text{-}3)$$

where,

Figure 7-3. Typical high pressure zone.



Roll	1	2	3	4	5	6	7
Diameter, in	36	30	24	21	13	10	8
L_1, in	94	16	15	16	11	17	10
L_2, in	16	15	16	11	17	10	24
Θ	244°	130°	114°	160°	185°	180°	180°
$\frac{\pi D \Theta}{360}$	76.6	34.0	23.9	29.3	21.0	15.7	12.6

psi_3 = average pressure on the sludge cake due to F_3
F_3 = lb of force due to belt elasticity per inch of belt width, where $F_3 = 2eE/D$
D = roller diameter, in
E = modulus of elasticity of the belt (i.e., stress/strain before yield point)
e = belt strain, Δ/L_1
Δ = belt stretch (L_0-L_i), cm or in
L_1 = tangent length of belt entering roller
L_0 = length of outer belt around roller between tangent points on adjacent rollers
L_i = length of inner belt around roller between tangent points on adjacent rollers

- Total pressure on the cake at any roller:

$$psi = psi_1 + psi_2 + psi_3 \qquad (7\text{-}4)$$

or

$$psi = 2[F_1 + F_2 + F_3]/D$$

With these equations the engineer can calculate the total pressure on the sludge cake at each roller to ensure that there is a gradually increasing pressure on each successive roller. These equations also allow determination of roller and shaft diameters and bearing size requirements, which can then be compared to the belt press manufacturer's specifications.

An example using these equations to evaluate a belt filter press design is shown below and is reproduced with permission of the author (4). The parameters needed to evaluate a design follow:

D = diameter of each roller, in
L_1 = tangent length of belt entering each roller
L_2 = tangent length of belt leaving each roller
θ = angle of wrap of the belt around the roller.

The above parameters are available from the manufacturer's specifications. Other parameters follow:

Q = sludge throughput rate, lb of solids/minute
t_A = cake thickness at the entrance to the high pressure zone, in
t_B = cake thickness at exit
C_A = cake solids concentration at the entrance to the high pressure zone as a decimal
C_B = cake solids concentration at the exit of the high pressure zone as a decimal

E = modulus of elasticity from belt manufacturer's specifications, lb/sq in
HP' = drive horsepower per inch of belt width per belt
W = belt width, in
t_1 = cake thickness at entrance of roller, in
t_2 = cake thickness at exit of roller, in
t_a = average cake thickness at roller, $(t_1 + t_2)/2$
L = effective length of belt (portion over each drum), Dθ/360.

For this example, the magnitudes of the values for D, L_1, L_2 and θ are shown in Figure 7-3. The magnitudes of the other parameters follow:

W = 80 in
Q = 125 lb/min
t_A = 4.0 in
C_A = 0.1
C_b = 0.38 (desired value)
HP' = 0.0127 HP/in
fpm = $2.3Q/(W \times t_A \times C_A)$ = 9.0 fpm

Procedure:

1. Calculation of cake thickness at exit of high pressure zone.

$$t_B = (0.95 \times t_A \times C_A)/C_B = 1.0 \text{ in} \qquad (7\text{-}5)$$

This equation assumes 95 percent solids capture across the unit.

2. Plot: $F_1 + F_2$, in lb/in vs. Dθ/360

3. Calculate:

$$e + \Delta/Li = (2t_a)/\{[360(L_1 + L_2)/\pi\theta] + D\} \qquad (7\text{-}6)$$

4. Calculate:

$$F_3 = 939e\text{-}[(100/2{,}000e^2) + 1] + 108 \qquad (7\text{-}7)$$

5. $\Sigma F = F_1 + F_2 + F_3$

6. psi = $2/(d\Sigma F)$

Comments and Interpretation of Results (Table 7-2):

1. The value of t_A will have some maximum allowable value according to the press design.

2. t_a is the value at the center of the roller, halfway between tangent points.

3. The total pressure (psi) calculated for each roller is compared to the pressures specified by the manufacturer. In this example, the sharp increase in pressure between rollers 4 and 5 suggests that sludge would either extrude into the belt mesh, clogging the belt or squeeze out from between the two belts. It is important to remember that the pressure should increase gradually from roller to roller and from zone to zone. Sharp increases of pressure can indicate operating problems.

4. In this example, the belt tension (ΣF) is greater than 250 lb/in in many cases. A force of this magnitude tends to deform most dewatering belts, which suggests problems with belt tracking and alignment as the belt wears.

5. The angle of wrap (θ) should be as close to 180° as possible.

6. Use F_1, F_2, and F_3 for analyzing the size of the rollers and bearings in the other zones. Failure to include F_3 when analyzing the high pressure section rollers and bearings could result in bearing and/or shaft failure.

7. The design engineer should require a submittal of the calculations from the manufacturer to confirm bearing and shaft design.

8. To calculate the reaction at each bearing for rollers in the high pressure zone with a bearing on each end of the roller, use Equation 7-8:

$$R = W[2(F_1 + F_2) + F_3] \sin(\theta/2) \qquad (7\text{-}8)$$

7.2.4 Mechanical Description

The mechanical components of a belt filter press generally include the following:

- Dewatering belts
- Rollers and bearings
- Belt tracking and tensioning system
- Controls and drives
- Belt washing system.

The dewatering belts are usually woven from monofilament polyester fibers. There are various weave combinations, air permeabilities and particle retension capabilities available from belt manufacturers. These parameters greatly influence how well the press will perform. For example, a sludge with a high concentration of activated sludge (80 - 100 percent) may require a belt with a high air permeability and high solids retention capability, while a primary sludge might require just the opposite. Therefore, since sludge types and qualities vary considerably from plant to plant, it can be important for the press manufacturers and engineers to try to evaluate different weaves, permeabilities, and solids retention capabilities for each installation to ensure optimum performance. The initial belt type is usually provided by the supplier based on the sludge type and his past experience.

Such an evaluation is very simple for a plant that is already in operation and is generating sludge that can be treated on a belt filter press. For a newly designed

Tablo 7-2. **Summary of Pressures Calculated for Each Roller**

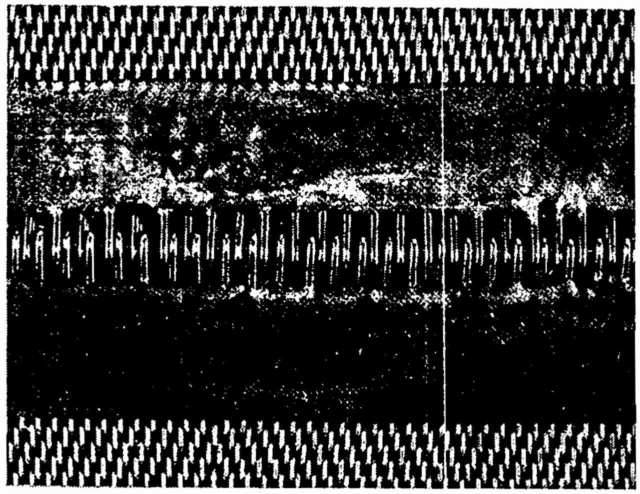
Roll	D	L_1	L_2	θ	(F_1+F_2)	t_i	t_2	t_a	e	F_3	ΣF	psi
1	36	94	16	244	43	4.0	2.9	3.45	0.079	174	217	12
2	30	16	15	130	53	2.9	2.4	2.65	0.092	188	241	16
3	24	15	16	114	58	2.4	2.1	2.25	0.082	178	236	20
4	21	16	11	160	63	2.1	1.7	1.90	0.094	190	253	24
5	13	11	17	185	67	1.7	1.4	1.55	0.102	197	264	41
6	10	17	10	180	70	1.4	1.2	1.30	0.096	192	262	52
7	8	10	24	180	73	1.2	1.0	1.10	0.074	169	242	61

plant, the belts must be evaluated differently. Sludge from a similarly designed plant can be tested or surveys can be made of similar plants to determine what type of belt they are using. Further, belt filter press manufacturers can supply information on how to relate belt characteristics to type of sludge expected. Usually, the belt filter press manufacturer can recommend the belt best suited to the type of sludge expected at the plant. Once the plant is operating, belts with different characteristics can be tried to obtain optimum performance. The belts should be designed for ease of replacement with a minimum of downtime to insure continuous dewatering. There are two different types of belts, split and continuous. The split belts are joined together with a splicing device called a clipper seam (see Figure 7-4). Split belts are the most common type on the market and can be used on all models of belt filter presses. The continuous or seamless belt can only be used on certain presses. The designer should consult the press manufacturer for a specific belt recommendation. The manufacturers of continuous belts claim longer life than with the split type. However, there is no available data to substantiate this claim. Continuous belts are more difficult to install.

The rollers and bearings are the main mechanical components of the belt filter press. As stated earlier, the rollers provide the pressure and forces that allow dewatering to occur, but they also insure proper belt support and tension. The tensioning device is the key control once the roller sizes have been fixed. Roller diameter and shaft sizes are key design parameters and should be carefully evaluated (Section 7.2.3). The bearings are extremely important components since they support and guide the rollers.

Press controls should be centralized either on the press itself or on a remote control panel (Figure 7-5) and should include automatic sequential start-up and shutdown systems, instrumentation for tracking and tensioning of the belts, pressure gauges, and safety interlocks. In addition, many engineers include running time meters, sludge and polymer pump controls, and other auxiliary devices that allow for a more efficient operation. The panels should be NEMA

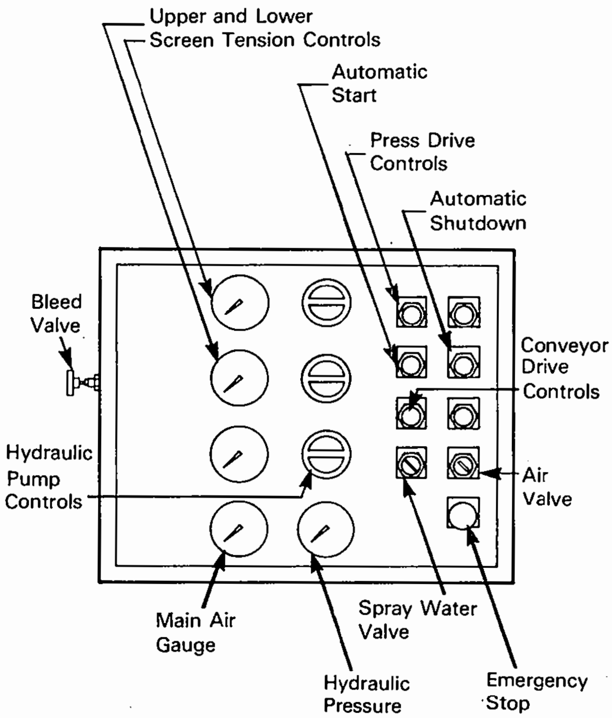
Figure 7-4. **Typical clipper seam for split dewatering belt.**



12 and should be well designed with centralized controls and adequate safety interlocks. If possible, they should be located in a separate control area away from noise, odors, and moisture that might affect the controls. Each piece of equipment which makes up the sludge dewatering system should be interconnected so that each unit is started in the proper sequence. For example, the press and conveyors should be started before the sludge and polymer feed pumps. In addition, automatic shutdown of equipment must be provided, also in the proper sequence. If a piece of equipment downstream fails, everything upstream of that unit must shut-down, i.e., if the press fails for any reason, the polymer and sludge feed pumps must shut-down automatically. Automatic shutdown of dewatering equipment should occur for any of the following fault conditions (3):

- Belt drive failure
- Sludge conditioning tank failure
- Belt misalignment
- Insufficient belt tension
- Loss of pneumatic or hydraulic system pressure
- Low belt wash water pressure
- Emergency stop

Figure 7-5. Control panel for belt filter press dewatering system.



- High sludge level on gravity drainage section
- Polymer feed pump shut-down (should stop feed pumps).

The belt washing system should include a high pressure water pump, a set of spray bars for cleaning both the upper and lower belts, and a spray cleaning device. Belt washing occurs after the cake has been removed from each belt. It is washed from the side opposite that which is in contact with the sludge cake. Either potable water or high quality plant effluent could be used as wash water for the presses. Since some belt presses require as much as 3.16 l/s (50 gpm) per meter belt width of wash water, it is more economical to use plant effluent. Therefore, when designing a system ensure that the pump, piping, and nozzles are capable of handling high quality plant effluent. However, when plant effluent is used, a highly efficient filtration system must be installed upstream of the press to ensure that the effluent is free of solids that can clog the spray nozzles. The spray nozzles should be designed for easy access to enable efficient and thorough cleaning which ensures complete cleaning of the filter belts. Many of the new models of presses are equipped with stainless steel brushes within the spray header to automatically clean the nozzles without removing them from the press. Figure 7-6 (5) illustrates this type of spray header. This type of cleaning system should be specified since they are easy to use and require much less operator time than the manually cleaned systems.

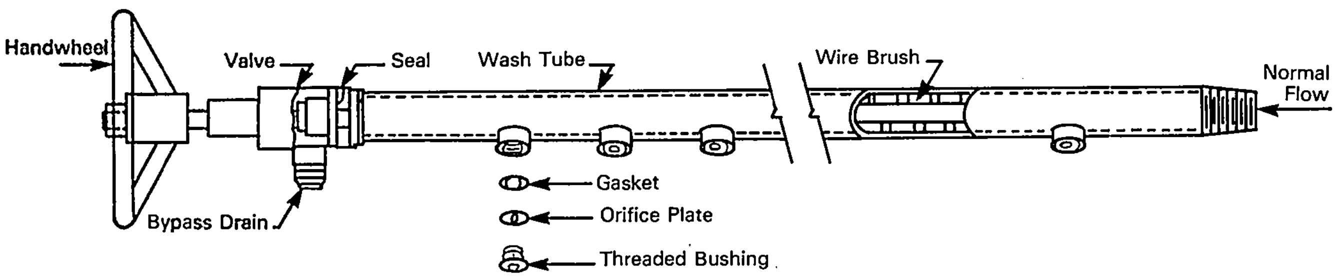
7.2.5 Performance Characteristics

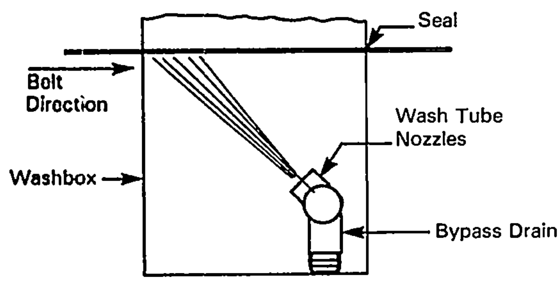
Belt filter presses can be used to dewater most sludges generated at municipal wastewater treatment plants. However, the sludge must be conditioned with polymer to ensure optimum performance. Polymer produces a phenomenon known as superflocculation (2). Superflocculation is the formation of large, strong floc which causes free water to drain easily from the sludge in the gravity drainage zone of the belt filter press. Superflocculation also produces a sludge that can withstand the pressures generated during the dewatering process and prevents the sludge from squeezing out from between the dewatering belts. Only polymer can produce this phenomenon. Some plants have tried to dewater lime conditioned primary sludge on belt filter presses. (H. Johnson, Ashbrook-Simon-Hartley, personal communication, 1987; J. Labunski. Parkson Corp., personal communication, 1987.) The results have been poor, with low cake solids, low solids capture, and blinding of the belt mesh. If the designer wants to use lime stabilization and landfilling as the final disposal method, a post-lime-stabilization system should be used. First, the sludge is conditioned with polymer and dewatered, and then lime is added to the sludge cake (6). New Haven, CT is successfully dewatering a high pH (lime added) raw primary and waste activated sludge using a cationic polymer. However, this can be site specific and results vary widely.

Usually cationic polymers are used for sludge conditioning. Sometimes a two polymer system, such as a cationic polymer following either an anionic or a nonionic polymer, will be used on a belt filter press to improve cake release from the upper dewatering belt. The polymer must be selected carefully to ensure optimum performance. Always contact the polymer manufacturer's representative for help screening and testing polymers; this service is normally provided free.

Typically, polymer and sludge are piped to the conditioning unit. When designing the polymer conditioning system, it is important for the designer to locate polymer feed points at several locations: one at the conditioning unit itself, one about 0.6 to 0.9 m (2 to 3 ft) upstream of the unit directly into the sludge feed piping, and one about 7.6 m (25 ft) upstream. With two polymer feeds, the anionic or non-ionic polymer may have to be added before the sludge pump. Feedpoint location is especially important for a new installation where sludge characteristics are not known, but is also important for any plant, since sludge characteristics can change periodically. Sometimes the sludge will condition better when there is a longer contact time with the polymer, but at other times it requires a shorter period. Therefore, by designing polymer feed points at several locations,

Figure 7-6. Washwater spray bar with cleaning brushes.





the designer can ensure flexibility and optimum performance. For more information on sludge conditioning, refer to Chapter 5. Typical dewatering data for various types of sludges is shown in Table 7-3 (3).

Typical polymer conditioning costs for belt filter press dewatering range from a low of \$2.65/Mg (\$2.41/dry ton) to a high of \$91.15 Mg (\$82.86/dry ton) with an average of \$24.38/Mg (\$22.17/dry ton).

Odors can be a problem during belt filter press operation. Odors can be controlled with good ventilation systems, ensuring that the sludge is kept fresh, and using chemicals such as potassium permanganate to neutralize the odor-causing chemical.

Potassium permanganate, $KMNO_4$, is a strong oxidizing agent which rapidly ozidizes the hydrogen sulfide in sludge to odorless sulfate. It is packaged as a dry, crystalline, water-soluble powder. The permanganate solution can be fed directly into the suction side of the sludge transfer pump so that the pump itself can act as a mixer for the permanganate and sludge. The permanganate feed point should be upstream of the polymer feed point by a distance which yields a travel time of about 1 minute (7) to ensure the most effective use of the $KMNO_4$. A typical installation is shown in Figure 7-7. It should also be noted that some plants use hydrogen peroxide for the same purpose as potassium permanganate.

A dosage of about 0.5-2.0 kg/Mg (1-4 lb/dry ton) is typical, but the dosage depends on the concentration of hydrogen sulfide in the sludge. Permanganate adds about \$1.00/Mg to the cost of dewatering the sludge but it produces several benefits besides the destruction of the sulfides. These include (7):

- Improved sludge dewaterability - slight increase in cake dryness using permanganate
- Reduction of polymer - increased sludge production with no additional polymer
- Sulfide-free recycle stream from the belt presses.

7.2.6 Design of a New Installation

This design example (8-10) is for a proposed 0.22-m^3/s (5-mgd) secondary, activated sludge, treatment plant with primary clarifiers. The plant will be dewatering raw sludge during an eight hour day, five days per week. The sludge will be thickened before dewatering to 5.0 percent solids. How many and what size presses will be required?

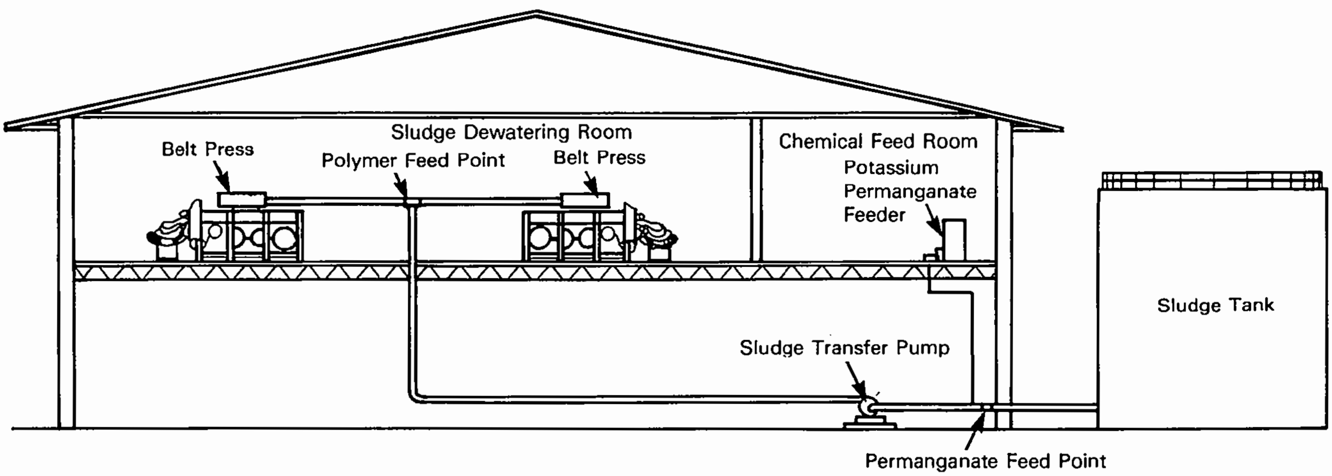
Belt presses are not sized on the basis of wastewater flow to the plant, but on the basis of the weight or volume of sludge to be dewatered. The following calculations show how the required number of presses can be determined.

- Determination of the amount of primary sludge: Influent total suspended solids concentration is determined by laboratory analysis (11) of the

Table 7-3. Typical Data for Various Types of Sludges Dewatered on Belt Filter Presses

Type of Sludge	Feed Solids	Solids Loading Rate	Polymer Dose	Cake Solids
	percent	kg/hr/m belt width	g/kg	percent
Raw:				
P	3-10	360-680	1-5	28-44
WAS	0.5-4	45-230	1-10	20-35
P+WAS	3-6	180-590	1-10	20-35
P+TF	3-6	180-590	2-8	20-40
Anaerobically Digested:				
P	3-10	360-590	1-5	25-36
WAS	3-4	40-135	2-10	12-22
P+WAS	3-9	180-680	2-8	18-44
Aerobically Digested:				
P+WAS	1-3	90-230	2-8	12-20
P+TF	4-8	135-230	2-8	12-30
Oxygen Activated:				
WAS	1-3	90-180	4-10	15-23
Thermally Conditioned:				
P+WAS	4-8	290-910	0	25-50

Figure 7-7. Belt filter press installation using permanganate for odor control.



wastewater that will be flowing through the plant. For this example, assume 220 mg/l (8).

Total suspended solids
= 220 mg/l x 5 mgd x 8.34 lb/Mgal/mg/l (7-9)
= 9,180 lb/d (4,168 kg/d)

A primary clarifier will remove an average of 60 percent of the suspended solids. Therefore:

Total primary sludge = 0.6 x 9,180 lb/d
= 5,508 lb/d (2,501 kg/d)

Assume the primary sludge is 3.5 percent solids.

Total volume of primary sludge
= (5,508 lb/d) ÷ [(0.035)(8.345 lb/gal)]
= 18,858 gal/d (71.38 m^3/d)

- Determination of the amount of waste activated sludge: Influent BOD_5 is determined by analyzing the wastewater (11). For this example, assume 220 mg/l (8), a 30 percent removal of BOD_5 in the primary clarifier, and a total plant removal of 90 percent of the BOD_5.

BOD Removal
= 0.7 (220 mg/l) - (0.1)(0.7)(220 mg/l) (7-10)
= 139 mg/l

BOD Removed
= 139 mg/l x 5mgd x 8.34 lb/Mgal/mg/l
= 5,780 lb/d (2,622 kg/d)

Assume a sludge yield coefficient of 0.5 @ SRT of 16 days:

Solids production = 0.5 (5,780 lb/d)
= 2.892 lb/d (1,313 kg/d)

Total waste activated sludge will then be equal to the amount of suspended solids in the primary effluent plus the solids production in the aeration system minus the effluent suspended solids.

WAS = 0.4 (9,180 lb/d) + 2,890 lb/d - 918 lb/d
= 5,646 lb/d (2,563 kg/d)

Assume that the waste activated sludge is 1 percent solids:

Total volume of WAS
= (5,646 lb/d)/[(0.01)(8.34 lb/gal)]
= 67,657 gal/d (256 m^3/d)

- Determination of the volume of the 5 percent thickened sludge to be dewatered per day.

The total volume of the 5-percent mixed sludge would be about 27,000 gpd.

The plant wants to dewater only 5 days per week, therefore:

(27,000 gpd x 7)/5 = 37,800 gpd (143 m^3/d) (7-11)

A typical belt filter press has a hydraulic loading of 40 gpm per meter of belt width:

37,800 gallons/40 gpm = 945 minutes = 16 hr

Therefore, a 1-m belt press would dewater the sludge produced at this plant in a 16-hr day. However, sludge is to be processed on an 8-hr day and it is also important to design for excess capacity. Therefore, two 1.5-m units should be used. With both units operating, the sludge would easily be dewatered in 8 hr. Further, there is enough excess capacity so that if one unit were out of service, the other unit could process all of the sludge in a 10.5-hr day, thus preventing a build-up of sludge in the plant.

Following is a summary of equipment and operating recommendations for the design of belt filter press installations:

A. Equipment
- Use durable materials for equipment construction.
- Provide sturdily constructed, properly coated frames.
- Use long-life bearings (L_{10} life of at least 100,000 hr)
- Use high strength rollers.
- Provide continuously acting tension/tracking systems.
- Use high quality, durable, and properly woven materials for belts.
- Ensure that both the press and belt manufacturers provide high levels of quality control.

B. Performance
- Consult manufacturers for design and performance data early in planning stage.
- Confirm performance data with other operating installations and/or through pilot testing.
- Specify high quality equipment and require a performance bond.
- Assure system integration by specifying that the dewatering system be the responsibility of a single supplier.

C. Auxiliary Equipment
- Provide sludge blending prior to dewatering to enhance continuity of feed sludge.
- Use a macerator upstream of the belt filter press to ensure a homogenous feed.
- Provide flexibility in points of polymer application and type of polymer used.
- Use continuously acting high pressure sludge pumps such as progressive cavity or rotary lobe.
- Provide positive ventilation for odor control in the dewatering area. Provide carbon or chemical odor control system.

D. Controls
- Provide instrumentation to monitor such operating parameters as sludge, filtrate, and washwater flow (including provisions for sampling).
- Integrate ancillary system controls with those for the dewatering equipment and interconnect key control functions.
- Protect controls from the moist, corrosive operating environment.

E. Safety
- Provide non-skid walkways and floors.
- Provide adequate access to equipment.
- Assure installation and maintenance of emergency stop systems, drive guards, and other protective equipment.
- Educate operators to follow safety precautions; assure adherence to rules.

F. Operations
- Monitor system performance to assure optimum operation.
- Assure that sludge is properly conditioned.
- Assure good gravity drainage of sludge.

G. Operator Training
- Provide operation and maintenance training upon completion of installation.
- Provide ongoing training to maintain skills.

7.3 Centrifuges

7.3.1 Introduction

Centrifugal dewatering of sludge is a process that uses the force developed by fast rotation of a cylindrical bowl to separate the sludge solids and liquid. In this basic process, when a sludge-water mixture enters the centrifuge, it is forced against the bowl's interior walls, forming a pool of liquid and sludge solids. 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 separately discharged from the unit. The two types of centrifuges used for municipal sludge dewatering, basket and solid bowl, both operate on these basic principles. They are differentiated by the method of sludge feed, magnitude of applied centrifugal force, method of solids and liquid discharge, cost, and performance. A third centrifuge type, the disc-nozzle centrifuge, has been used for thickening waste activated sludge (WAS), but does not produce a dewatered material. It will not be discussed in this manual.

Engineers have long recognized the potential of centrifugal dewatering devices for handling both domestic and industrial waste slurries and sludges. Recent improvements in materials, design, and process technology have not only produced sturdier, more sophisticated centrifuges, but also have given rise to new operational procedures and design practices. These developments have alleviated or eliminated past problems, and have increased process performance levels.

A five-decade evolution preceded the centrifuge's acceptance in municipal wastewater treatment. Though simple in concept, the practical aspects of centrifuge design are quite involved and are still controversial. Following the 1902 development and testing of the first perforated basket centrifuge in Cologne, Germany, Herman Schaefer joined forces with Dr. Gustav Ter Mer in 1907 to produce improved basket units. These batch units were installed in Germany and the USSR. A modified Schaefer-Ter Mer unit was installed during 1920 in Milwaukee, Wisconsin to thicken activated sludge. It could not produce clear centrate at an economical rate and this finding was further verified during the twenties and thirties.

A solid-bowl conveyor centrifuge with continuous feed and discharge of solids was developed and tested during the thirties for dewatering of raw and digested sludges. Two decades later an improved version was successfully installed and operated in California. It marked the start of a gradual acceptance of the solid bowl centrifuge by the wastewater treatment industry. This unit's bowl was 1.5 times as long as its diameter; produced a centrifugal force in the range of 900 to 1,500 Gs; and produced cake solid concentrations of 20-35 percent with a solids recovery of 50 to 70 percent. An improved solid bowl centrifuge, producing a force of over 3,000 Gs and with a bowl to length diameter ratio of 2.5:1, was introduced in the early 1960s. It had better recovery of sludge solids and higher cake solids. In 1965 a German unit was produced that had a bowl length diameter ratio of 2.9:1 and operated at forces in the range of 2,500 to 4,000 Gs. While it gave even better performance, wear was a significant factor.

In the 1970s, significant improvements were made in centrifugal designs and sludge conditioning. In the early 1970s German manufacturers developed a slow speed concurrent centrifuge, which could produce high recoveries and acceptable cake solids; they introduced the use of the hydraulic backdrive arrangement. This backdrive device provided automatic control of the centrifuge bowl and conveyor differential speed, using the change in the conveyor torque output to adjust differential speed to maintain constant torque. In 1974, a major advancement in protection of wearing surfaces was the development of special hardened, easily replaceable tiles on the wearing surfaces. Four- to eight-fold increases in the operating life of the conveyors resulted.

From 1974 to 1986, there were other refinements in construction of centrifuges to enhance performance and reduce O & M costs. Machine sizes and capacity grew with sludge centrifuges available up to 183 cm diameter (72 in) x 427 cm (168 in) bowl length and with capacities exceeding 37.9 l/s (600 gpm). These machines are used in both thickening and dewatering of sludges. Both high-speed and low-speed centrifuges are employed for all types of sludge.

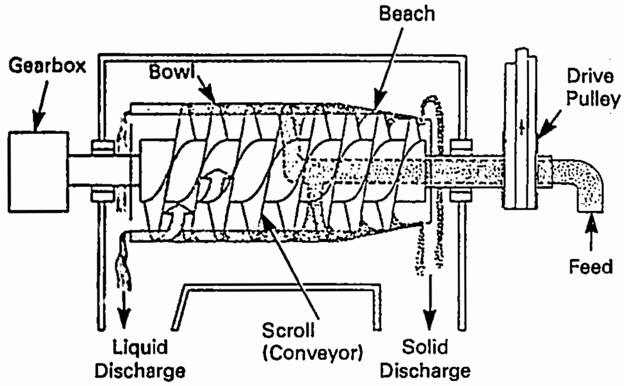
As the advancements in solid-bowl centrifuges have primarily been occurring in the conveyor type, there has been a declining interest in the basket centrifuge for municipal sludge dewatering.

7.3.2 General Centrifuge Type and Description

The solid-bowl centrifuge, also called decanter, conveyor, or scroll centrifuge, is characterized by a rotating cylindrical-conical bowl. A helical screw conveyor fits inside the bowl with a small clearance between its outer edge and the inner surface of the bowl. The conveyor rotates, but at a slightly lower or higher speed than the bowl. This difference in revolutions per minute (rpm) between the bowl and the scroll is known as the differential speed, which allows the solids to be conveyed from the zone of the stationary feed pipe, where the sludge enters, to the dewatering beach, where the sludge cake is discharged. As shown in Figure 7-8, the scroll pushes the collected solids along the bowl wall and up the dewatering beach, located at the tapered end of the bowl, for final dewatering and discharge.

The differential speed between the bowl and conveyor is maintained by several methods. Earlier designs used a double output gearbox that imparted different speeds as a function of the gear ratio. It was possible to vary the output ratio by driving two separate input shafts. Eddy current brakes are also used to control the differential. Latest designs provide automatic speed control as a function of conveyor torque that can maximize solids concentration.

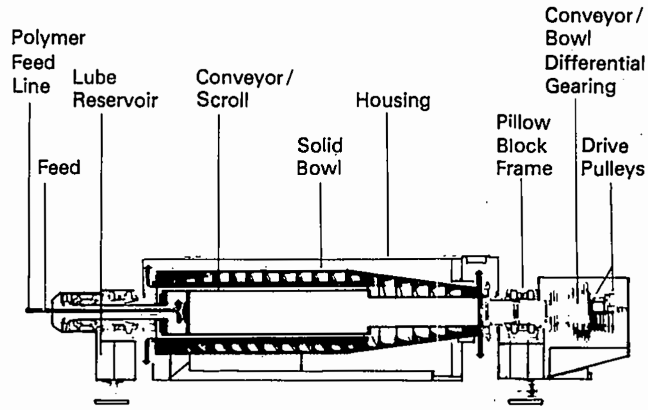
Figure 7-8. Solid-bowl (countercurrent) conveyor discharge centrifuge.



The solid-bowl centrifuge operates in one of the two modes: countercurrent or continuous concurrent. The major differences in design pertain to the location of the sludge feed ports, the removal of centrate, and the internal flow patterns of the liquid/solids phases. In the countercurrent centrifuge, influent sludge is introduced through the feed pipe at or near the junction of the cylindrical and conical sections of the bowl; the solids move to the conical end of the machine while the centrate flows in the opposite direction. This design is shown in Figure 7-8. Under the influence of centrifugal force, the sludge solids are pushed against the bowl wall. The solids are then moved gradually by the rotating conveyor along the bowl wall, up the dewatering beach. From there, they drop into a sludge cake discharge hopper. Centrate, the partially clarified liquid containing smaller and some finer unflocculated solids, flows around (and through) the conveyor toward the liquid discharge end. Depending on the sludge particle characteristics, gravitational forces and residence time, a portion of these solids settle to the outer wall and are also conveyed to the solids discharge end while the liquid flows over the adjustable weir. The length of the conical section (drying beach) above the pool level may vary considerably depending on the specific sludge characteristics and the centrifuge design/operation. In general, the lower the structural strength of the cake, the smaller the desirable drying beach length.

In the concurrent model, shown in Figure 7-9, feed slurry is introduced at the opening opposite the dewatering beach. The settling zone then begins near or at the feed point, and the solids travel the full length of the bowl. While general construction is similar to the countercurrent design, the centrate flows in the same direction as the sludge solids (concurrent flow) and is withdrawn by a skimming device or return tube located near the junction of the cylindrical bowl and the conical section. Clarified centrate then flows into channels inside the scroll hub and returns to the feed end of the machine, where it is discharged over adjustable weir plates through outlet ports built into the bowl head.

Figure 7-9. Solid-bowl (concurrent) centrifuge with hydraulic scroll drive.



The concurrent design of conveyor centrifuges is most often operated at speeds lower than the countercurrent design and in the range of 700 to 1,500 gravities depending on machine size and sludge properties. It has been categorized as a "low-speed" or low-G centrifuge, which operates at 50-75 percent of the gravitational force of a high-G centrifuge.

7.3.3 Applications

A solid-bowl centrifuge's ability to be used either for thickening or dewatering provides flexibility and is a major advantage. For example, a centrifuge can be used to thicken ahead of a filter press, reducing chemical use and increasing solids throughput. During periods of downtime of the filter press, the solid-bowl centrifuge can serve as an alternate dewatering device. Another advantage for larger plants is the solid bowl centrifuge's sludge throughput capability, which allows the largest single units of any type of dewatering equipment. The larger centrifuges are capable of handling 19 to 38 l/s (300 to 600 gpm) per unit depending on the sludge's characteristics. The centrifuge also has the ability to handle higher than design loadings, such as a temporary increase in hydraulic loading or solids concentration, and the percent solids recovery can usually be maintained with the addition of a higher polymer dosage. The cake solids concentration will likely decrease, but the centrifuge will handle the higher solids loading.

As discussed in Chapter 3 for all sludge dewatering processes, it is helpful to at least partially thicken the feed to the centrifuge, so that capacity is not limited due to the excessive water content of the sludge. Ideally, the feed solids are sufficiently pre-concentrated such that the centrifuge liquid (Sigma) and solids (Beta) capacity are reached at the same time. Diluted feeds can result in reduced solids capacity as well as increased polymer requirements.

High rate preconcentration of dilute sludges in a gravity thickener using loadings that are 200-400 percent higher than normal gravity thickener loadings can improve performance, reduce costs due to partial thickening of the sludge, and provide a more stable operation using the reservoir of sludge in the thickener. Recommendations for gravity thickener loadings and underflow concentrations for both conventional and high rate are provided in Table 7-4.

Table 7-4. Conventional Gravity Thickening and High Rate Pre-Concentration

	Conventional		High Rate	
Sludge	Loading	Underflow	Loading	Underflow
	$kg/m^2/d$	percent	$kg/m^2/d$	percent
Raw P	98	8-10	196	5-6
Raw WAS	20	1.75 + SDI	74	0.5 + SDI
(PS + WAS) (60:40)	40	4.5-5.5	98	3.5-4.0

SDI = Sludge Density Index.

7.3.4 General Design Theory and Considerations

The solid-bowl centrifuge is essentially a high energy (g) settling unit. Particles entering the liquid pool settle toward the outer wall aided by gravity and resisted by the same factors that slow settling and resist compaction in clarifiers and thickeners. The capacity of the centrifuge is also affected by the type of solids, rate of solids removal, solids concentration, etc. Figure 7-10 compares the clarifier and centrifuge. Whereas the clarifier is large, operating at 1 G, the centrifuge clarifier-thickener operates at high multiples of gravity to separate liquid and solids in a shallow, small volume unit. The high gravitational force permits a greater removal of water, resulting in a dewatering function.

Although there are some major deficiencies, the settling velocity of a particle in a fluid under the influence of gravity can be defined by Stokes Law as follows:

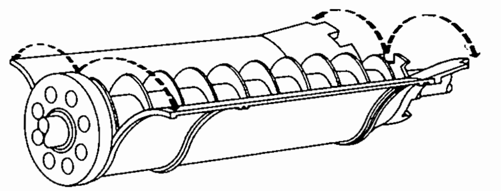
$$V = g\,(\rho_{solid} - \rho_{liquid})d^2 \div 1{,}800\mu \qquad (7\text{-}12)$$

where,

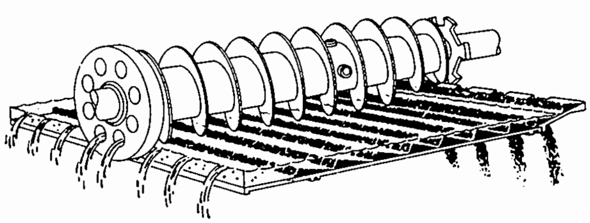
- V = settling velocity of solid particle in the fluid, m/s
- g = acceleration due to gravity, m/s^2
- ρ_{solid} = particle density, kg/m^3
- ρ_{liquid} = liquid density, kg/m^3
- d = mean particle diameter, m
- μ = viscosity of liquid, kg/(m x s)

For settling in a centrifuge, the same relationship can be used but with g, the acceleration of gravity,

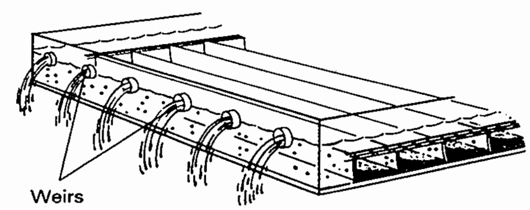
Figure 7-10. Comparison between the clarifier and the centrifuge (courtesy Pennwalt Corp., Sharples-Stokes Div.)



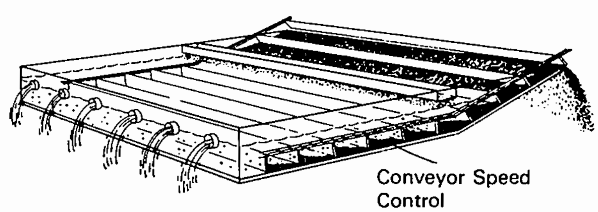
A centrifuge has the same basic characteristics as a clarifier. It is a clarifier that has been wrapped around a center line so that it can be rotated to generate g's.



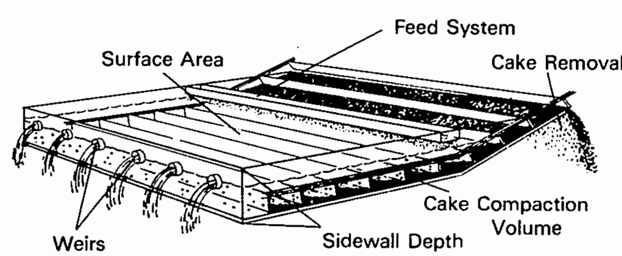
By looking at the design of a centrifuge as a clarifier, several design improvements can be incorporated.



For example, redesigned overflow weirs reduce material turbulence at the liquid overflow.



The installation of an eddy current brake controls conveyor speed (i.e., cake removal rate) and makes full use of the solids compaction volume.



In sum, the centrifuge's basic design elements, which are like a clarifier's, can be refined to take full advantage of surface area, detention time, weir design, and other factors.

replaced with G, the centrifugal force produced on the particle by the rotation of the bowl.

The centrifugal acceleration force (G) defined as multiples of gravity is a function of the rotational speed of the bowl and the distance of the particle from the axis of rotation. In the centrifuge, the acceleration force, G, is calculated as follows:

$$G = (2\pi N)^2 R \div 60 \qquad (7\text{-}13)$$

where,

- N = rotational speed of centrifuge, rev/s
- R = radius of rotating body of liquid, m

The value of G can be substituted for g in the earlier equation to determine the rate of settling velocity as a function of R. This theory presents the basis to argue for a very shallow liquid film for settling. In practice, this must be modified by sludge compaction requirements, turbulence zones, clearance requirements between solids and clarified liquid zones, etc., such that medium to deep pools are now used for sewage sludges. Heavier solids, like $CaCO_3$, can be efficiently dewatered using shallower pool depths since they readily compact and are easily conveyed (scrolled) out of the machine.

The centrifuge has three functions that are not entirely compatible. The first is clarification or removal of solids from the liquid suspension, and the second is consolidation of the settled particles against the bowl wall. Lastly, it is necessary to convey and further dewater these solids during their transport out of the bowl. At a specific gravitational force, the effectiveness of clarification and solids concentration will each be a function of the pool volume, a large volume favoring good clarification and a small volume favoring high solids concentration. That is, detention volume available for clarification will affect the hydraulic rate, separation efficiency and, perhaps, the chemical dosage. However, maximizing the clarification volume (minimize sludge volume, hence sludge depth) reduces the time for the solids to compact and a lower sludge concentration must be the result. Generally, the operating mode is a compromise between objectives. If scrolling (moving) of the solids is difficult because the movement needs to resuspend the particles, it may be necessary to compromise both the clarification rate (reduce feed rate) and solids concentration (increase pool depth).

The clarification capacity of a solid-bowl centrifuge has historically been measured by its Sigma (Σ) value as defined by Ambler with certain assumptions (12). The Sigma value is essentially the averaged surface area of a settling tank equivalent to the sedimentation capacity of the centrifuge, and it is given by the following formula:

$$\Sigma = 2\pi L (w^2/g) (0.75 r_1^2 + 0.25 r_2^2) \qquad (7\text{-}14)$$

where,

- Σ = theoretical hydraulic capacity, m^2
- L = effective clarifying length of centrifuge bowl, m (inlet to liquid outlet)
- w = angular velocity of centrifuge bowl, rad/s
- g = acceleration due to gravity, 9.8 m/s^2
- r_2 = radius from centrifuge centerline to the liquid surface in the centrifuge bowl, m
- r_1 = radius from centrifuge centerline to the inside wall of the centrifuge bowl, m

A simplified method of calculating Sigma that is applicable only to a solid-bowl centrifuge is given by the equation:

$$\Sigma = P_v w^2/g \, [\ln (r_2/r_1)] \qquad (7\text{-}15)$$

where,

P_v = the pool volume, m^3

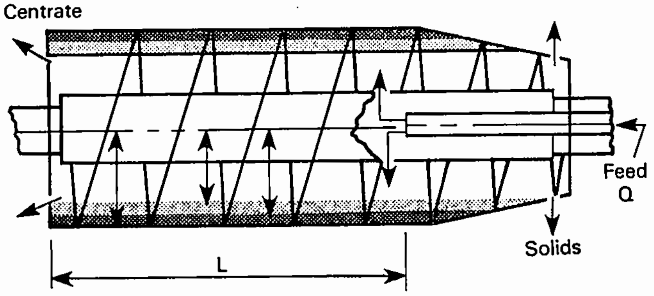
The Sigma value can be used to estimate relative performance characteristics between centrifuges but has limitations (12). It can be and is used for scaling up of results from machines of comparable operating conditions and physical configuration. However, the use of Sigma to compare a high-G centrifuge capability to a low-G capability (or ones where relative pool depths are markedly different) can be and often is invalid. The scale-up also may be modified by the manufacturers' experiences with the specific machines involved.

Scale-up of a smaller test machine to a larger, similar solid-bowl centrifuge using Sigma is shown in Figure 7-11. The use of Sigma is based on an application where the centrifuge is only clarification limited and solids capacity is unquestioned. Clarification capacity can generally be enhanced using chemicals to help flocculate and settle the particles. The effectiveness of polymers and the greatly increased clarification area (longer bowls) have generally resulted in the sludge consolidation -- transport being the limiting process rate factor. This is particularly true when high solids concentrations are desired or required for a cost/effective sludge handling operation. Sigma can be the limiting factor when handling dilute feeds.

The solids limiting capacity of a solid bowl centrifuge has been designated as the Beta value, and it is used like Sigma. Solids capacity of a centrifuge is reached at the point where the compacted sludge volume in the pool interferes with the solids-liquid separation and solids recovery declines. The total pool volume (V) in a centrifuge is equal to (see Figure 7-11):

$$V = \pi (r_1^2 - r_2^2) L \qquad (7\text{-}16)$$

Figure 7-11. Sigma scale-up procedure.



$$\Sigma = 2\pi L \frac{\omega^2}{g} (0.75r_1^2 + 0.25r_2^2)$$

Test Centrifuge A	Test Centrifuge B
L = 0.861 m	L = 1.625 m
r_1 = 0.178 m	r_1 = 0.318 m
r_2 = 0.143 m	r_2 = 0.254 m
ω = 366 rad/s	ω = 262 rad/s
Σ_A = 2,134 m²	Σ_B = 6,574 m²
Q_A = 9 m³/hr	Q_B = ?

$$Q_B = \frac{\Sigma_B}{\Sigma_A} Q_A$$
$$= 27.7 \text{ m}^3/\text{hr}$$

The volume of sludge, V_s, in the pool with a surface at r_3 is:

$$V_s = \pi (r_1^2 - r_3^2) L \qquad (7\text{-}17)$$

The solids flow into a centrifuge is defined as Q_s kg TSS/hr, and the volumetric flow of the solids by Q_s/Y_c, where Y_c is the specific weight of the compacted solids, kg TSS/m³ in the cylindrical portion of the bowl. If there is no slippage of solids in the bowl, then the particle travel time (T) can be calculated from a formula for the conveyor or scroll as:

$$T = L/\Delta w \; SN \qquad (7\text{-}18)$$

where,

- L = length of cylinder, m
- Δw = differential speed, rad/s (speed at which solids are conveyed out of centrifuge)
- S = spacing between conveyor blades, m
- N = number of leads

The volume occupied by solids during steady state operation is:

$$V_s = (Q_s/Y_c)\, T$$
or
$$V_s = (Q_s/Y_c)\,(L/\Delta w \; SN) \qquad (7\text{-}19)$$

Since the surface area of the inside wall is $A = 2\pi r_1 L$, then the depth of the cake, y, is:

$$y = (V_s/A) - [Q_s/Y_c \; 2\Delta w \pi r_1 \; SN] \qquad (7\text{-}20)$$

$$V_s/V = (r_1^2 - r_3^2) \div (r_1^2 - r_2^2)$$
$$= (Q_s/Y_c) \div 2\Delta w \; SN \; r_1 \; (r_2 - r_1)$$

where,

β = 2 Δw SN r_1 $(r_1 - r_2)$ and is expressed as m³/hr of cake conveyed from the machine.

Now let's look at how Σ and β can be used in making design-related decisions. Referring to Figure 7-11, if r_3 for test centrifuge A was equal to 0.163 m, then the value of V and V_s would be as follows:

$$V = \pi (0.178^2 - 0.143^2)(0.861) \qquad (7\text{-}21)$$
$$= 0.0304 \text{ m}^3$$

$$V_s = \pi (0.178^2 - 0.163^2)(0.861)$$
$$= 0.0138 \text{ m}^3$$

If the test machine A was solids limited, indicated by a deterioration in centrate quality, at 9 m³/hr feed rate, then the larger machine is assumed to be solids limited at the same ratio or:

$$(V_s/V_A) = (V_s/V_B) \qquad (7\text{-}22)$$

The pool volume in the larger machine is 0.187 m³. Therefore, the equivalent ratio of sludge volume in centrifuge B would be:

$$V_s = (0.0138 \text{ m}^3)(0.187 \text{ m}^3) \div 0.0304 \text{ m}^3 \qquad (7\text{-}23)$$

The Sigma and Beta scale-up factors determined in this example are:

	Centrifuge A	Centrifuge B
Sigma - m²	2,134	6,574
Flow - m³/hr	9	27.7
Beta Ratio	1	2.43
Flow - m³/hr	9	21.9

While the clarification rate (Sigma) indicated a potential scale-up capacity of 27.7 m³/hr, the maximum capacity of the centrifuge is only 21.9 m³/hr due to solids flux (Beta). However, it would be possible to operate the centrifuge at a higher differential speed and increase solids capacity and sacrifice dryness of the cake.

7.3.4.1 Laboratory Testing/Field Evaluations
Bench-scale tests have limited use and should not be considered a replacement for continuous laboratory or field evaluations. Laboratory testing, which requires the shipment of large quantities of sludge and encounters subsequent deterioration of

the sludge with time, can provide misleading results. It has been reported (13) that the use of formaldehyde at a dosage of 5 mg/l prevented changes in the dewatering characteristics of the sludge.

Where small, continuous field-scale tests are practical, there are some limitations in terms of magnitude of the scale-up. It is recommended that the scale-up factor using Beta and Sigma not exceed 3 whenever possible.

Since there are substantial differences in the design approach to the load chambers of high-G machines, it is suggested that comparative field tests provide the most useful information for design. This is particularly true when the cake solids concentration could have a significant impact on the economics of downstream operations. If tests involving two machines are contemplated, they should be run concurrently. Side-by-side operation will alleviate any concern that the sludge characteristics changed during the tests between the two machines. It is further recommended that the test machines be similar to the full-scale units planned. The use of torque controlled back-drives, for example, should be included if that is the plan for the full-sized machines. The testing range of flow and solids rates tested should be adequate to provide a full description of the operating characteristics of the machine. Ideally, these tests should also be conducted during the colder months of the year, since that will be the most difficult time to dewater the sludge. If it is done during warmer periods, then the cake solids should be discounted to account for cold weather operation.

Since the feed solids are split between the centrate and the cake, it is necessary to use a recovery formula to determine solids capture. Recovery is the mass of solids in the cake divided by the mass of solids in the feed. If solids contents of the feed, centrate and cake are measured, it is possible to calculate percent recovery without determining total mass of any of the streams. The equation for percent recovery is given below:

$$R = 100\,(C_S/F)\,[(F - C_C)/(C_S - C_C)] \qquad (7\text{-}24)$$

where,

R = recovery, % TSS
C_S = cake solids, % TSS (or TS)
F = feed solids, % TSS
C_C = centrate or overflow solids, % TSS

7.3.5 Centrifuge Components, Operation and Control

While not specifically part of the centrifuge, the foundation upon which it rests is an important design consideration. The base provides a solid foundation on which to mount and support the centrifugal unit. Vibration isolators, normally mounted between the base and foundation, help reduce the vibration created by the centrifuge. The base is normally fabricated steel or cast steel of sufficient mass to sustain vibration and reduce harmonic effects caused by minor imbalance.

The centrifuge's case serves as a guard, protecting the rotating assembly and reducing the noise level. It also contains and directs the cake solids and centrate as they are discharged from the rotating assembly. The case may be fabricated from carbon steel and coated, but the cake and centrate discharge housings should be of stainless steel of SS316 quality or better.

The variables listed below will be discussed in this section. Many are preset by the manufacturer; some can be controlled by the operator.

- Bowl diameter
- Bowl length
- Bowl rotational speed
- Beach angle
- Beach length
- Pool depth
- Scroll rotational speed
- Scroll pitch
- Feed point of sludge
- Feed point of chemicals.
- Condition of scroll blades

7.3.5.1 Bowl and Conveyor

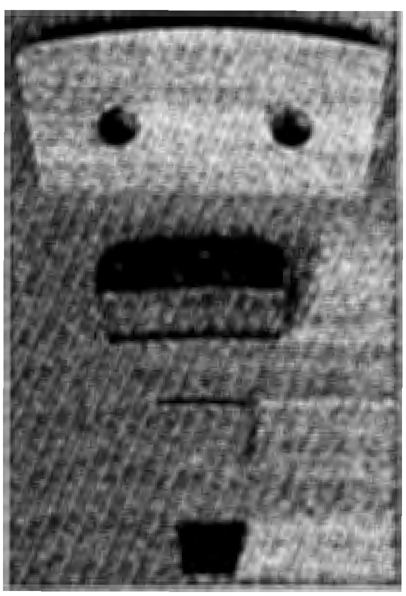
The bowl configuration is cylindrical and conical in shape, though the proportions of each section will vary depending on manufacturer and application. The angle and length of the conical section, which acts as the dewatering beach, have an important affect on the performance for individual applications and will be specific to each manufacturer's machine. The bowl diameters of dewatering centrifuges range up to 183 cm (72 in) with bowl lengths of up to 427 cm (168 in) in 1986. In 1975, the largest units in the United States were 91 cm (36 in) diameter x 244 cm (96 in) long.

The scroll, a helical screw conveyor, is fitted concentrically into the bowl. The central core of the scroll contains feed tubes and ports for the discharge of the centrate. Design of the scroll may vary considerably in terms of the pitch diameter, number of conveyor leads, and in openings for the centrate to pass through to the discharge weirs. The scroll may contain baffles to prevent the incoming feed from disturbing the previously consolidated cake in a countercurrent centrifuge. Improved conveyor designs are often jealously guarded secrets and the final decision regarding conveyor details must be left to the manufacturer.

The material of construction for the centrifuge bowl and conveyor ranges from high strength carbon steel to common stainless steels such as SS316 and SS317. Special alloys are used for some applications, but normally this is not required for municipal sludges. Where there is a concentration of chlorides above 200 mg/l, there is concern for chloride stress corrosion of stainless steels. Plants receiving chloride wastes or infiltration of sea water may have serious mechanical problems if this is not considered.

Abrasive wear on scroll conveyor blades or flights has traditionally been the item of greatest maintenance. It is influenced by sludge abrasiveness, the centrifugal force at the bowl wall, the differential speed, and the abrasion resistance of the material used to form scroll blade tips. Figure 7-12 shows the various types of hardfacing that have been used to reduce wear on scroll tips. These include many different welder applied metallic hardfacings (such as Colmonoy #6, Eutalloy, and Stellite) as well as tungsten carbide and ceramic tiles. Field replaceable ceramic tiles have recently been recommended by low-G centrifuge manufacturers because of their long life, relatively low replacement cost, and ease of replacement. However, they are more fragile than metallic hardfacings, tending to chip easily. They also may occupy more space in the bowl and do not form as smooth a surface on the conveyor blades as do metallic hard facings. Ceramic tiles can be glued onto the flights although in some cases they are both glued and bolted to the flights. One manufacturer of low-G centrifuges using ceramic tile hardsurfacing material will routinely guarantee scroll conveyor life for 15,000 to 20,000 hours between rebuilds.

Figure 7-12. Four different types of hardfacing used to retard scroll wear.



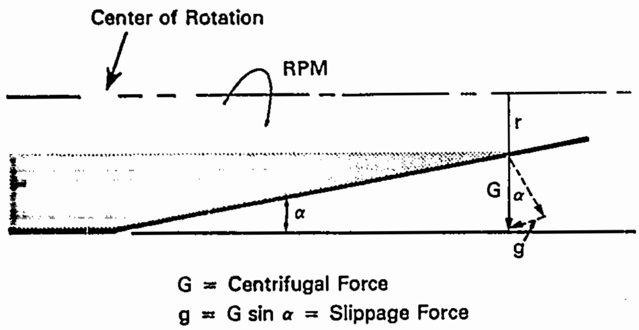
The Severn Trent Water Authority in Birmingham, England reported in 1981 that the original ceramic tile conveyor lasted 25,000 hours and the replacement units had operated 27,000 hours and were still satisfactory (T. Wood, Severn Trent Water Authority, Birmingham, England, personal communication, 1981). The machine averaged 165 hours per week operation and was a low-G centrifuge. Similar experiences with ceramic and sintered tungsten carbide tiles have been reported in the United States at San Francisco, Oakland, CA, Port Huron, MI, and Lorain, OH.

Sintered tungsten carbide tiles have demonstrated useful lives greater than 30,000 hours, but they are generally more expensive than ceramic tiles. However, the cost may decrease as additional suppliers and refurbishing plants employ these materials. Sintered tungsten carbide tiles are generally welded to the flights and are usually required for only the portion of the conveyor blade near the dewatering beach. One high-G centrifuge manufacturer warrants scroll conveyor life for 30,000 hours using highly abrasion resistant sintered tungsten carbide tiles. Experience with low-G concurrent flow centrifuges at the Los Angeles County Sanitation District's Carson Plant has indicated that conventional welder applied hardfacing has an operating life of only 5,000 hours.

Bowl and scroll geometry varies considerably from one manufacturer to another. In general, increasing the bowl diameter will increase both the capacity of solids conveying and clarification. On the other hand, an increase in the bowl length improves only clarification capacity.

The beach angle is usually kept at 8 to 10 degrees to help prevent slippage of the conveyed solids. As the solids emerge from the pool, the buoyancy effect is lost and it becomes more difficult to convey fine, hydrous, and soft solids, such as waste activated sludge against the high G forces of the centrifuge. Shallow beach angles, deep pools, and conveyor design configurations also work together with a hydraulic effect, which essentially helps push the settled solids up the beach and eliminates slippage problems. Although a shallow beach angle increases conveyor capacity and improves centrate quality, a wetter cake is produced due to the loss of beach drainage area. Conversely, a steep beach angle produces a drier cake but at the expense of centrate quality and conveyor capacity. With a steep beach angle, there are higher resistive forces to conveyance as shown in Figure 7-13. Albertson and Guidi (14) reported that the force of slippage can increase 10-fold at the pool-beach interface. As a result, a portion of the solids are generally resuspended or leaked through the beach and are ultimately lost over the centrate weir. The ultimate decision of the beach angle depends on the relative importance of greater cake solids or centrate quality.

Figure 7-13. Effect of bowl angle on the movement of sludge.



Bowl speed is normally not varied on most centrifuge models once the unit is installed. The solid-bowl centrifuge operates at speeds equivalent to 600-3,000 times the force of gravity and are categorized into low- and high-G centrifuges. Low-G units have operating speeds equivalent to 600-1,800 Gs, and high-G units operate at 2,000-3,000 Gs. The gravitational force is directly proportional to the bowl diameter and the square of the bowl speed. Thus, since the G force takes into account both bowl speed and bowl diameter, it is a better method of describing solid-bowl centrifuges than bowl speed alone.

The question of which type of centrifuge, high-speed or low-speed, performs best cannot be answered generally. The results have been site and sludge specific; higher gravitational forces have produced slightly poorer to significantly better cake solids and recovery. Sludges with higher structural characteristics, as with a paper fiber content, will generally respond well to a higher gravitational force, while weak structured cakes may even respond negatively to increased gravitational forces.

Arguments for the low-speed, concurrent centrifuges have been as follows:

- Feed introduction at far end of bowl reduces turbulence, improves clarification.
- Feed introduction does not disturb partially consolidated sludge.
- Chemical consumption is lower due to less turbulence.
- High solids and clear centrate can be produced at lower speeds without loss of machine capacity.
- Lower speed requires less power.
- Lower speed reduces wear and other O & M costs.
- Lower speed produces less vibration.

Arguments for high-speed, countercurrent centrifuges have been as follows:

- Speed can always be reduced if not required.
- Most municipal sludges are not tested prior to installation of the centrifuge to determine the performance characteristics as a function of speed or G force.
- Concurrent centrifuges will wear the full length of the bowl and scroll, whereas countercurrent units wear only a portion of the bowl, resulting in less repair costs per repair interval.

In general, the high-speed, countercurrent centrifuges will be smaller in size than the concurrent, low-speed unit of similar hydraulic and solids capacity. Some operating/design principles also differ, and further, it is not always possible to reduce the gravitational force since the higher speed would be required to maintain clarification capacity. Without testing, some of these considerations will not be readily apparent.

There are specific benefits to both design concepts that often can be determined only by side-by-side evaluations. A basic rule to follow is: operate at the minimum speed possible that still meets the capacity and other performance characteristics necessary for a cost-effective dewatering system.

The City of San Francisco has both high-G (HSC) and low-G (LSC) centrifuges dewatering anaerobically digested primary and pure oxygen activated sludge (G. Davies, City of San Francisco, personal communication, 1983; J. Loiacono, City of San Francisco, personal communication, 1986). Also, there is an existing vacuum filter (VF) station. Average results reported by the City are provided in Table 7-5. Additional results and comments regarding operation and performance are found in Chapter 9.

Table 7-5. Sludge Dewatering Performance at San Francisco (1982-1983)

	HSC	LSC	VF
Feed, l/s	7.32	8.64	-
TSS, kg/hr	634	740	-
Cake, % TS	23.6	22.7	16.2
Recovery, % TS	96.9	96.7	88
Polymer, kg/Mg	4.1	3.6	-
Power, kWh/m^3	1.9	1.1	-
Gravities, G	1,880	1,000	-

HSC - 74 cm diameter x 208 cm L (29 in x 62 in)
LSC - 90 cm diameter x 225 cm L (35 in x 96 in)

The difference in cake solids was negligible particularly when one considers that the LSC was handling 17 percent more solids and operating at a lower polymer dosage. At San Francisco, the solids content of the cake is not strongly affected by the gravitational force.

Extensive field trials (15) were conducted at the Littleton/Englewood STP, Colorado with a high-speed and a low-speed centrifuge for thickening WAS (TWAS) and dewatering digested primary and waste activated sludge [D(P+WAS)]. The results of the tests are summarized in Table 7-6.

The HSC was able to thicken the WAS to 7 percent TSS without polymer addition while the LSC required polymer at a cost of $4.40/Mg DS to achieve 7.0 percent TSS at a higher recovery. The results on the digested sludge were very similar. However, it was necessary to operate the LSC at higher than the customary G force of 1,000-1,400 to achieve equivalent performance. This sludge was amenable to the use of higher G forces to maximize performance.

Centrifuges require a stationary feed pipe, which is inserted through the center of the scroll housing for a distance necessary to reach the discharge port. The feed pipe may enter the centrate or the cake discharge end of the centrifuge depending on the machine size and design.

Polymer conditioning of sludge could still be termed an art. Thus, it is important to have maximum flexibility for polymer addition. Provisions for adding polymer before and after the feed pumps, centrifuge inlet and into the centrifuge feed discharge ports through an independent polymer feed line should be provided as part of the design. See Chapter 5 and Appendix B for greater discussion of Chemical Conditioning.

7.3.5.2 Differential Bowl and Conveyor Speed Assembly

The gear unit is generally of a planetary or cyclo-gear type and works together with the backdrive to control the differential speed between the bowl and the conveyor. By controlling the differential speed, optimum solids residence time in the centrifuge and the cake solids content can be provided. A backdrive of some type is considered essential when dewatering raw or digested primary - secondary or secondary sludges due to the presence of fine particles. The backdrive function can be accomplished with a hydraulic pump system, an eddy current brake, DC variable speed motor, or a Reeves type variable speed motor. The two most common backdrive systems are the hydraulic backdrive and the eddy current brake.

The control of the bowl-conveyor speed differential has had a series of evolutionary changes over the past 25 years. The early centrifuges were provided with a fixed gear ratio, which, in turn, fixed the ratio of the bowl speed to the conveyor speed. An improvement was the use of an auxiliary motor, which controlled the speed of the previously fixed output shaft of the gearbox. The eddy current brake backdrive is now provided on high-G centrifuges. The eddy current brake is attached to the pinion shaft of the gearbox and consists of a stationary field coil and a brake rotor on the shaft. When a DC voltage is applied to the stationary field coil, magnetic flux lines are created in the brake rotor. The amount of flux in the rotor is a function of the speed differential between the rotor and the field coil as well as the DC current applied to the field coil. This flux produces eddy currents, which create a resistance to turning, or a braking action. Thus, varying the DC voltage applied to the stationary field coil will change the speed differential between the bowl and the scroll. While the eddy current backdrive differential can be easily set by the operator, it is still possible to overload the conveyor and cause blockage of the centrifuge due to overtorque.

The most versatile backdrive arrangement is a hydraulic pump design. This arrangement is widely used since it completely eliminates the need for a gearbox and a mechanical or electrical backdrive. The hydraulic unit is now employed by many centrifuge suppliers since it can assure that the centrifuge conveyor is not overloaded and can maintain the bowl-conveyor differential at the optimum level. A unit manufactured in Switzerland is used by many centrifuge manufacturers to provide automatic differential control. It has a low-speed hydraulic motor that drives the centrifuge scroll independently of the bowl. A pump unit powers the hydraulic motor and its control system senses the scroll torque and regulates the differential speed to prevent blockage. The hydraulic scroll speed control can operate at lower differential speeds than the fixed gearbox differential. As a result, an increase in cake dryness and recovery will be possible. Lower differential speeds also reduce the scroll tip speed, which, in turn, mitigates the wear on both the conveyor and the bowl shell or facing strips. Since any change in solids loading, hence torque, will be automatically compensated for by a change in differential speed, the operation of the centrifuge can be tuned for maximum retention of solids within the bowl and scroll, without the risk of choking the machine. That is, the differential speed is increased (or decreased) in proportion to an equivalent change in torque. Furthermore, increased throughput is possible, since the automatic torque-related scroll speed controller can allow the feed rate to be increased without danger of plugging.

An all-hydraulic system eliminates the problem of gearbox failure and another problem, torsional vibration or chattering, is minimized. Reliability and improved performance are the major reasons that the all-hydraulic system is widely employed. Since the scroll is directly driven by the adjustable volume hydraulic system, it also eliminates the need for a separate backdrive/eddy current brake system, a common maintenance problem. As the feed solids to the machine change, the scroll backpressure is

Table 7-6. Centrifuge Tests at Littleton/Englewood STP, Colorado

	HSC		LSC	
	TWAS	D (P+WAS)	TWAS	D (P+WAS)
Feed, l/s	3.15	3.79	3.15	3.79
Cake, % TS	7	18	7	17
Recovery, % TSS	75	90	85	95
Polymer, $/Mg DS	0	14.55	4.40	14.55
Power, kWh/Mg	364	472	273	364
Gravities, G	2,500	2,500	1,185	1,800
Operating Costs, $/Mg	15.45	20.00	13.64	18.18

HSC - 42.5 cm diameter x 125.7 cm L (16.75 in x 49.5 in)
LSC - 45.7 cm diameter x 134.6 cm L (18 in x 53 in)

continuously monitored and the analog hydraulic unit automatically increases or decreases the scroll speed according to pressure. Because the scroll differential speed is independent of the bowl speed, there is no loss of torque in the event that the main drive motor shuts down. In contrast, the gearbox-driven centrifuge must contend with torque- related problems. Once the torque limit is reached in a gear-driven centrifuge, either a shear pin breaks, a clutch disengages, or a current relay trips, thus causing loss of torque and differential speed, and increasing the risk of the machine plugging.

The hydraulic oil pressure is a direct measure of the torque; as such, the torque loading on the scroll can be read at any time and can also be used for feed rate control. The control unit can also be used to control polymer dosage based on the clarity of the centrate. A representation of the one unit's electronic monitoring box is pictured in Figure 7-14.

Controlling Bowl/Scroll Differential Speed to Optimize Performance

The differential speed controls the solids residence time within the centrifuge and thus, it can greatly influence cake concentration and machine capacity. Increasing the differential between the bowl speed and the scroll speed normally results in a wetter sludge cake and higher machine throughput. Conversely, a decrease in the differential speed produces a drier cake and decreased machine throughput and may result in poorer recovery or higher polymer dosage to maintain recovery. One of the problems of operating at too low a differential speed is the creation of a pile of solids in front of the scroll conveyor blades, allowing some of the fine solids to be skimmed from the top of the cake into the centrate. Another danger is plugging the centrifuge if solids are removed at a slower rate than they are fed to the machine.

A backdrive unit can generally provide an increase in cake solids content of 4 percent or more relative to a comparable machine without a backdrive. The

Figure 7-14. Bowl/scroll differential speed monitoring box.

backdrive increases the overall stability of the centrifuge performance when the feed solids characteristics vary. Retrofitting a centrifuge with an analog hydraulic backdrive is cost-effective. At Seattle Metro wastewater treatment plant, the existing centrifuges were modified to provide torque controlled conveyor differential speed. The cake solids increased 3-5 percentage points and the estimated savings in haulage were $39,000 per month. The test results of a study conducted by Seattle Metro (16) are shown in Table 7-7. As shown in the Seattle studies, the polymer cost can increase when operating at the minimum differential speed. The lower differential speed produced auto torque control, the sludge inventory was higher, and more polymer was used to control effluent solids. The cost of

Table 7-7. Performance of Fixed vs. Variable Torque Controlled Backdrive at Seattle Metro

Differential Control	Feed Rate	Cake TS	Recovery	Liquid Polymer
	m^3/hr	percent	percent TSS	kg/Mg
Fixed Differential	50.8	16	80	103
Automatic Backdrive	50.8	20	81	106
Fixed Differential	45.2	16	85	62
Automatic Backdrive	45.2	21	86	77
Fixed Differential	37.5	16	85	45
Automatic Backdrive	37.5	19	86	72

Haulage Savings @ 50.8 m^3/hr = $39,000/month.

haulage, fuel, etc., must be balanced against chemical cost.

Tests were also conducted on a 74 cm (29 in) dia. x 208 cm (82 in) long centrifuge in Columbus, OH. Providing torque directed conveyor speed control manually produced a cake that was 4-7 percentage points drier than other units on-line which were handling the same feed solids. Similar results have been experienced elsewhere.

7.3.5.3 Miscellaneous

Overflow Weirs

Although the pool depth is variable on solid-bowl units, several hours of labor may be required to adjust the overflow weirs. As such, it is not a popular method of operational control. The pool depth regulates both the quality of clarification and the dryness of solids. Thus, while increasing the pool depth will normally result in better solids recovery at a specific feed rate, the cake produced will be wetter. However, it may be necessary to adjust the weirs if a major change in feed rate is required. When the pool depth is translated to residence time in the centrifuge, there may be little or no difference in the recovery and cake solids. However, this is not necessarily true for all types of waste sludges, nor for pool depths that leave little or no dry beach. Over the years, pool volumes have increased to accommodate the greater space occupied by solids under longer retention times. Additional depth and greater cross-sectional area reduce turbulence and permit solids to become compacted without interfering with the clarification process. The use of deep pool operation in thickening and dewatering waste activated sludge minimizes the slippage force on the beach, resulting in improved conveying efficiency.

Sludge Feed Pumps and Piping

Control of sludge feed rate demands a sludge pumping system that can handle varying sludge consistencies and centrifuge loadings. For this reason, progressive cavity pumps are the overwhelming preference of centrifuge designers. Lobe pumps also provide a steady flow, are positive displacement, and thus suitable for centrifuge feed pumps. Centrifugal pumps, on the other hand, are less adaptable to changes in sludge consistencies. Varying sludges affect the pumping rate and, as such, appropriate flowmeters and controllers - and possibly variable speed drives - are recommended for positive control of centrifuge loading.

7.3.6 *Process Variables*

In addition to machine variables, there are a number of process variables that affect the performance of a centrifuge. These variables are listed below.

- Feed rate
- Sludge characteristics
- Particle size
- Particle shape
- Rheology
- Solids concentration
- Liquid viscosity
- Liquid density
- Temperature
- Type of chemicals added.
- Amount of chemicals added

7.3.6.1 Feed Rate

One of the most important control variables during centrifuge operation, as already noted, is the feed rate of the centrifuge, both from a hydraulic and solids loading standpoint. The hydraulic load to the centrifuge affects the clarification ability, while the solids loading is a function of the conveying capabilities. Increasing the hydraulic load will decrease the centrate clarity and may increase the chemical consumption. A corresponding change in the differential speed is required when changes in solids loading occur if the centrifuge is initially operating at maximum solids residence time. The most concentrated cake is achieved at minimum differential speed and at a feed rate to match the reduced volumetric conveying capacity.

7.3.6.2 Sludge Characteristics

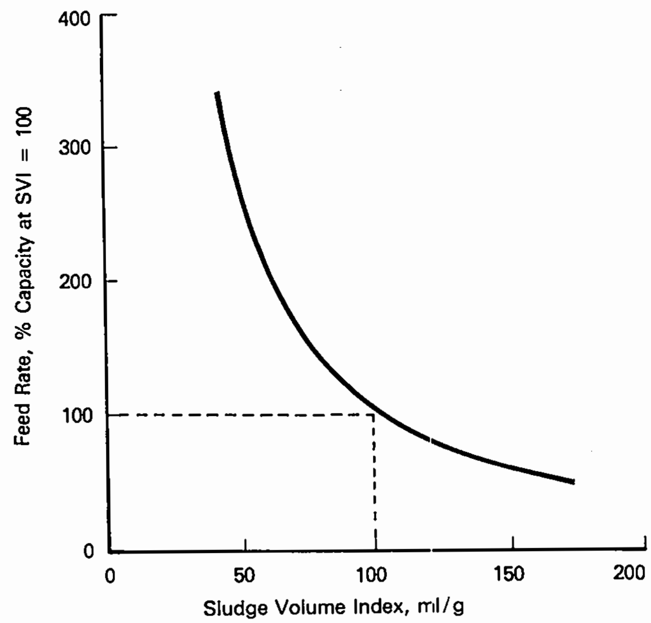
The identifiable characteristics of the various sources of waste solids have an impact on the dewatering efficiency as measured by unit capacity, product dryness, and solids recovery. Thus, the design of the wastewater treatment plant is an important consideration in the sizing of the centrifuge and the expected performance characteristics.

Larger and heavier particles are most easily captured by the centrifuge. Finer particles that cannot be settled separately must be agglomerated by chemicals to a size that will settle in the pool to the outer wall, where they can then be conveyed to the discharge point. As the proportion of finer particles increases, the sludge becomes more difficult to flocculate and requires increasingly higher dosages of chemicals to maintain a high capture of the feed TSS. Also, as the proportion of finer particles increases, the cake moisture content will also increase. If the finer particles are also hydrous, as is the case with activated sludge or alum sludge, the moisture content can increase significantly. The sludge cake produced will also change characteristics; it will have much more of a thixotropic nature as the proportion of fine and hydrous particles increases. Also, the sludge will be more plastic when the WAS fraction increases.

The sludges with a high proportion of fine and hydrous particles will also have poor structural characteristics. That is, the solids will have a tendency to flow. This characteristic will affect the conveying of solids from the centrifuge. As the sludge becomes more fluid, it will resist being conveyed up the slope of the conical portion of the bowl to the discharge point. If the upward frictional force of the conveyor is less than the centrifugal force on the solids, the solids may slip back into the pool.

The ash content of a sludge affects the final cake solids. Generally, about the first 10-15 percent of the inerts in the sludge are associated with the organics and thus have little impact. However, as the ash content exceeds 25 percent, a definite improvement in cake solids is noted. The added inerts will usually be fine silt, which dewaters readily and thus produces a higher cake solids. When comparing operating data, the designer must evaluate the possible effects of inert content. The Sludge Volume Index (SVI) of the secondary sludge can also have a profound impact on both the feed rate as well as the dewatered cake concentration. Vesilind and Loehr (17) found that the centrifuge capacity was impacted by SVI and their results are shown in Figure 7-15. The higher values of SVI will also produce a wetter cake.

Figure 7-15. Effect of SVI on centrifugal dewatering of activated sludge.



7.3.6.3 Chemical Conditioning.

As is described in detail in Chapter 5 and Appendix B, both inorganic and organic chemicals are used for dewatering applications. For the most part, solid-bowl centrifuges use organic polyelectrolytes for flocculating purposes. Polymer use improves centrate clarity, increases capacity, often improves the conveying characteristics of the solids being discharged and often increases cake dryness. Anionic polymers may yield a better operation if aluminum or ferric salts are present. A number of centrifuge installations are using $FeCl_3$ in conjunction with cationic and anionic polymers. It may be necessary to use a dual-polymer system if there is polymer treatment upstream of the centrifuge.

7.3.6.4 Sludge Temperature

Warm sludges will dewater better than cold sludge. Winter to summer sludge cake concentrations may vary by as much as 2-4 percentage points. The probable reason for the improvement in the summer cake is the decrease in liquid viscosity, which improves the liquid drainage. Another reason for the drier cake, though perhaps to a lesser degree, is the decrease in liquid density during warmer temperatures.

Secondary sludge quantities and proportions increase in winter, which, in turn, increase the moisture. Heating of the sludge will significantly improve the cake solids. However, this practice is rarely found to be feasible, unless there is an available source of usable waste heat, generally steam. One pound of steam is needed to heat 10 pounds of sludge to about 66°C (150°F).

7.3.6.5 Performance Characteristics

Primary to secondary ratio will have a profound effect not only on the capacity of the centrifuge, but also on the cake concentration and the polymer dosage. Further, as mentioned elsewhere, the SVI of the secondary fraction will also have an impact on these same parameters. Even if the sludge is digested, some of the effects of the primary to secondary sludge ratio and the SVI appear to carry through, affecting the centrifuge performance as well as other mechanical dewatering equipment. Wherever possible, the plant design should be directed toward producing a minimal amount of secondary sludge. This strategy will enhance the performance of the centrifuge as well as reduce the operating costs. It is difficult to compare data from different locations unless the primary and secondary sludge ratio and the SVI characteristics of the secondary fraction are known. Further, addition of chemicals for phosphate removal complicates this comparison and may make any conclusion invalid. The performance characteristics of a solid-bowl conveyor centrifuge on various sludges are provided in Table 7-8.

Mixtures of sludges can be estimated on a weighted mass basis; this is assuming that each fraction dewaters proportionately to its weight in the mixture. The procedure is as follows:

RPS	100 kg @ 30% TSS	=333 kg
RWAS	80 kg @ 16% TSS	= 500 kg
$Al(OH)_3 + AlPO_4$	30 kg @ 14% TSS	= 214 kg
	210 kg	1,047 kg

Cake TS = 210/1,047 = 20.1% TS

Polymer requirements can be determined in the same manner. The higher cake concentrations are generally achieved with the more favorable ratio of primary sludge. With the chemical sludges resulting from phosphorus removal, the range of solids can be due to a number of factors not well understood. The higher ratio of hydroxide precipitates will tend to reduce the solids content and part of the cause is generally an increasing fraction of secondary solids.

Centrifuge cake solids can be increased (2-5 percentage points) by using excessive polymer dosages, i.e., dosages above that necessary for 90-95 percent TSS recovery. Polymer costs are the lowest when the machine is running at a reduced capacity. Maximizing the capacity of the centrifuge will not only increase the polymer cost but also produce a wetter cake.

7.3.7 Pre-treatment

The sludge, prior to being pumped to the centrifuge, should be ground into a particle size in the range of 0.64 cm (1/4 in) or smaller. This is particularly true for the units in smaller treatment plants, which may have relatively small openings for feed inlets and discharge. In the very large plants, grinding of the sludge may not be necessary.

While materials of construction for abrasion resistance have greatly improved, good grit removal should be incorporated into the plant design. At a minimum, plus 65 mesh grit should be removed at peak flows entering the plant. This means that at normal flows, removal of plus 100 mesh should be readily achieved.

The best operation of the centrifuge will be achieved if the feed rate of solids loadings is relatively steady. In section 7.3.3, the benefits of preconcentration of the sludge using high rate thickening was discussed. The result of minimizing fluctuations will generally be higher solids concentration, better recovery, and lower polymer dosages. Operator attendance will also be minimized.

7.3.8 Cake Solids and Centrate Handling

Depending upon the application of the sludge characteristics and composition, the cake structural characteristics will vary widely. The cake may vary from a wet, sloppy mass to a relatively dry, firm solid mass, and in some cases, the solids will be a loose bulked product. Due to the high energy prior to when the sludge solids leave the machine, the solids will generally be massive, even if drier than a vacuum filter cake, for example. This is due to the thixotropic nature of the sludge and the conditions under which it exited the centrifuge. Thixotropic solids lose their structural integrity when energy, such as vibration, is applied. These solids are normally conveyed from the centrifuge by belt conveyors, screw conveyors, or specially designed pumps. The pump method has gained popularity since it is a very clean means of moving the sludge solids a considerable distance. Specially designed progressive cavity pumps, employing separate feeder mechanisms, have been used to pump centrifuge cake up to 61 m (200 ft), depending on the sludge concentration. More recently, ram type pumps, which are capable of pumping the cake greater distances, are being employed. These methods are not only effective; they are also less costly and much easier to maintain than the conventional conveying mechanisms.

Under normal operating conditions utilizing polymers, the centrate solids will constitute 5-8 percent of the feed solids. These solids are normally recycled back to the head of the plant or to a concentration unit prior to the centrifuge. These solids should not constitute any significant additional load on the clarification devices in the main plant stream. There will be a BOD associated with the centrate, both from the solids fraction as well as the soluble BOD contained in the liquid. It is recommended that the calculations of the feed sludge to the centrifuge include a factor of at least 10 percent by volume increase to account for a recycled load.

The centrate piping, in general, must be of adequate size and slope to allow for air venting the windage generated by the centrifuge. The manufacturer should be consulted for recommendations and the representative should review and approve the final drawings before bidding or installing the centrifuge. The centrate piping must also be sufficiently large to handle a mixture of foam and water. Often when polymers are employed, a stable foam can be produced. If the piping is too small, liquid will back up into the centrifuge. Liquid backup can be a problem particularly if there are multiple centrifuges

Table 7-8. **Centrifuge Performance Characteristics**

Sludge	Cake TS	Solids Recovery	Polymer	
	percent	percent	kg/Mg	$/Mg*
Raw Primary	28-34	90-95	1-2	5-10
Anaerobically Digested Primary	26-32	90-95	2-3	10-15
Raw WAS	14-18	90-95	6-10	30-50
Anaerobically Digested WAS	14-18	90-95	6-10	30-50
Raw (Primary + WAS)	18-25	90-95	3-7	15-35
Anaerobically Digested (P + WAS)	17-24	90-95	3-8	15-40
$Al(OH)_3 + AlPO_4$	12-16	90-95	1-3	5-15
$Fe(OH)_3 + FePO_4$	12-16	90-95	1-3	5-15
Ext. Aeration or Aer. Digested Sludge	12-16	90-95	6-10	30-50
$Ca_x(PO4)_x$	12-18	90-95	1.5-3	7-15
$CaCO_3$	40-50	90-95	0	0

* Based on $5/Mg. 40:60 to 60:40 P:WAS mixtures.

discharging into one centrate line. Here again, the manufacturer should be consulted during review for approval of the engineering design.

7.3.9 *General Equipment Selection Criteria*

The criteria used for selecting a centrifuge will be different depending on the plant size. That is, the criteria for smaller plants will emphasize reliability without complex servicing and maintenance while performance would be secondary. The generalized major (M) and secondary (S) criteria have been set forth in Table 7-9.

Table 7-9. **Evaluation Criteria for Centrifugal Dewatering Equipment**

	Plant Size (mgd)			
	Small (<2)	Med. (2-10)	Large (>10)	Cent. Rating
Quick Startup	M	MS	S	1
Reliability w/o Skilled Service	M	MS	S	2
Minimal Operator Attendance	M	MS	S	1
Parts Available Overnight	M	M	S	2
Local Repairability	M	MS	S	2/3
Cleanliness	M	M	M	1
Low Noise Level	M	M	MS	2
Low Initial Cost	M	MS	MS	2
Maximum Cake Solids	S	MS	M	2
Recovery Solids >90%	M	M	M	1
Polymer Cost	S	SM	M	1
Power Cost	S	SM	M	3
Labor Cost	M	M	M	1

M = Major; MS = Significant; S = Secondary (Minor)
1 = Excellent; 2 = Good; 3 = Fair

Small plants generally have low staffing levels, which are best served by unit operations that require minimal operator attendance. Large plants can use one operator to attend several machines and thus the labor cost per ton can be relatively small. On the other hand, driest possible cake could be very necessary for an economical combustion or compost operation. In this case, a higher capital and O&M cost for power, polymer, and labor could be offset by fuel savings and increased capability of the combustion or compost operations.

7.3.9.1 Capacity

It is not easy to compare capacities and performance of centrifuges of different manufacturers. As already noted, when low-speed centrifuges are compared to high-speed ones of similar capacity, they will be larger in diameter. While a centrifuge supplier will have a database that can be utilized to determine the capacity of these machines, independent analysis of operating units is recommended.

Appendix C includes tables that represent the bowl diameters and lengths of various suppliers. With some units, it is necessary that the engineer be assured that the selected design unit is still being manufactured. Further, new units of different sizes and configurations would normally be added to a manufacturer's line.

The capacity of sludge dewatering to be installed at a given plant is a function of the size of the plant, capability to repair malfunctioning machines on-site or locally, and the availability of an alternative disposal means. Proportionally, a smaller plant will have a higher percentage of standby capacity than a larger plant. On the other hand, the larger plant may have two or three spare machines while the small plant has one spare machine and an acceptable and available alternative disposal method. Some general guidelines relating the minimal capacity requirements are incorporated into Table 7-10. This table is based on the assumption that there is no alternative mode of sludge disposal and that the capacity to store solids is limited. These considerations may vary from case to case and must be considered individually by the design engineer.

Table 7-10. Suggested Capacity and Number of Centrifuges

Plant Size	Sludge Flow*	Operation	Centrifuges Operating + Spare
mgd	m^3/d	hr/d	@ m^3/hr
2	40	7	1 + 1 @ 6
5	50	7.5	1 + 1 @ 12
20	320	15	2 + 1 @ 12
50	800	22	2 + 1 @ 18
100	1,600	22	3 + 2 @ 25
250	4,000	22	4 + 2 @ 45

* Sludge production is about 0.1 kg/m^3 (1,600 lb TSS/mgd).
mgd x 0.0438 = m^3/s.

7.3.9.2 High-G vs. Low-G Centrifuge

For the medium to larger size plants, use of higher gravity rotational forces may increase the solids content as well as the capacity of a centrifuge. However, this is not true for all applications, and it is desirable to determine the ability of the dewatering operation to load to the high G force range. In any case, the operation of the centrifuge should be evaluated at different gravitational forces to ensure that the minimum G forces are employed for the specific application. This will extend the operating life of the machine. The question of whether high G forces are better than low G forces for dewatering cannot be answered easily. Solids with high structural strength respond more favorably to high G forces. Conversely, if the sludge contains substantial quantities of fines or hydrous materials, such as alum sludges and waste activated sludge, then low G forces can be equally effective. Only side-by-side tests can determine the comparative results.

7.3.9.3 Differential Speed Control

All centrifuges should be specified with an easily adjustable, differential speed control device. One option is the eddy current brake, which provides a readily adjustable fixed speed between the bowl and the conveyor. A better device is the hydraulic torque controlled device, which eliminates the maintenance associated with a gearbox and provides automatic, optimum control of the differential speed.

It is generally recommended that, for even the small machines, an automatic torque-controlled backdrive be employed. This device will optimize the machine's performance and produce the best results possible in terms of cake solids and recovery. Also, this type of drive is efficient, compensating for the normal variations found in small plants. The drive's ability to eliminate centrifuge overloading is sufficient reason in itself to install one. The machine will shut down if the bowl and scroll lock together and, at that point, the feed pump is automatically turned off.

7.3.9.4 Chemical Feed Control

Electronic monitoring of the torque and control of the speed differential comprise the basic equipment required to translate centrate clarity into chemical feed pump control. For example, as the clarity decreases, as measured by light transmittance, the polymer dosage will automatically be increased to maintain the preset clarity. If the chemical dosage is excessive, the torque setting is manually reduced to lower the sludge level in the pool. Since centrate feed control will provide optimal polymer dosing, save chemicals, and reduce operator attendance, this device should be incorporated into both large and small dewatering stations.

While there are devices available that can give a readout of solids concentration in mg/l, a light scattering absorption unit connected to the centrate line will probably provide more than adequate information regarding the changes in the recovery rate of the centrifuge. This continuous readout of the turbidity as a function of polymer doses will help to improve the dewatering operation as well as provide a record of the chemical consumption.

7.3.9.5 Abrasion Resistance

While there are applications where normal hardfacing has provided good life, the average wastewater treatment sludge is abrasive and, in some cases, highly abrasive. For this reason, it is recommended that the ceramic or sintered carbide tiles be used for sewage sludge dewatering units. These tiles will give a normal life of 15,000-30,000 hr and are suitable for the level of maintenance normally found in wastewater treatment plants.

The manufacturer will also recommend that other portions of the centrifuge be tiled in order to extend the life of the entire machine.

7.3.9.6 Materials of Construction

Experience has shown that both carbon steel and stainless steel will provide satisfactory service. Stainless steel machines can be more easily disassembled than carbon steel units after several years of service. However, tight tolerances on stainless steel may also be difficult to separate since the metal tends to grab. Chloride resistance is a consideration where sea water or higher than normal chlorides are present in the wastewater. Stainless steels are usually unsatisfactory for chloride exposure.

The use of stainless steel or carbon, to a large extent, depends on the engineer's and user's preferences. There is no strong technical reason to support either material since both materials provide satisfactory long-term service.

7.3.10 Performance Data for Solid Bowl Centrifuges

Table 7-11 includes operating data for a number of centrifuge installations that were installed since 1980. The results vary widely, which may be due to the machine design, solids loading, sludge feed characteristics, planned operating results, or a combination of these factors.

Suppliers have advised that there is limited data on newer machines that have been installed in the period from 1983 to the present. Since there has been continued advancement in the area of centrifuge design, the consulting engineer should carefully evaluate the new designs and the performance of these units. Any questions should be resolved by on-site testing.

7.4 Filter Presses

7.4.1 Introduction

Filter presses for dewatering were first developed for industrial applications and, until the development of diaphragm presses, were only slightly modified for municipal applications.

The original or early models of the press were sometimes called plate and frame filters, because they consisted of alternative frames and plates on which filter media rests or are secured. The frames provide both structural integrity and spacing between the plates. The frames could be changed to provide different cake thicknesses. The unit had a fixed and a movable end, which promoted pressure maintenance during the filtration cycle. There are few, if any, plate and frame units in service for municipal applications today, because this configuration is not particularly suitable for the filtration of hydrous pseudoplastic materials like municipal sludges. However, test filters sometimes are plate and frame, and they are used to determine the optimum cake thickness.

The equipment commonly in use for the dewatering of municipal sludges falls into one of two categories. The fixed-volume recessed plate filter and the diaphragm filter press; the latter was introduced within the last ten years.

A typical fixed-volume recessed plate filter press is shown in Figure 7-16.

Pressure filtration whether it be in a recessed plate filter press or in the diaphragm filter press is defined as confined expression. Precoating the filter media or substantial chemical conditioning of the sludge is normally required. This is particularly true for such difficult-to-dewater materials as waste activated sludge or aerobically digested sludge. A chemical conditioning station is therefore almost invariably a part of a facility that uses filter presses to dewater municipal sludges.

There is usually a higher degree of operator activity associated with filter presses than with most other types of dewatering. As a result, filter presses, for the most part, have been employed in wastewater treatment facilities $\geq$1.1-2.2 m^3/s (25-50 mgd). On the other hand, this equipment can produce a very dry cake, probably the driest cake produced by conventional dewatering equipment. Hence, it has substantial attractiveness.

7.4.2 Equipment Description

The recessed plate filter press shown in Figure 7-16 consists of a series of plates, each with a recessed section that forms the volume into which the sludge is pumped for dewatering. Filter media are placed against each wall and retain the sludge solids while permitting passage of the filtrate. The surface under the filter media is specifically designed to facilitate the passage of the filtrate while holding the filter cloth. Sludge is pumped with high pressure pumps into the volume between the two plates and individual pieces of filter media. The filtrate passes through the cake and the filter media and out of the press through special ports on the filtrate side of the media.

The pumping of sludge into the press continues up to pressures sometimes in excess of 1,380 kPa (200 psi). When solids and water fill the void volume between the filter cloths and ultimately no further filtrate flow occurs, pumping is stopped. Shortly thereafter, the press is opened mechanically, and the cake is removed. Practice has separated the operation of recessed plate filters into two principal categories: low pressure units and high pressure units. Low pressure units operate between 350-864 kPa (50-125 psi) as the terminal pressure; high pressure units operate between 1,040-1,730 kPa (150-250 psi). Typically, the low pressure units will terminate at about 691 kPa (100 psi) and the high pressure units at about 1,380 kPa (200 psi). There are several ways of maximizing the filtrate removal, including good conditioning and stepping the pressure. Stepping is particularly effective for the high pressure units, using increments of 350-520 kPa (50-75 psi).

The diaphragm press is a comparatively new device, having been commercialized in the U.S. in the 1980s. It operates during the initial filling period, if the sludge is properly conditioned, very much like a gravity drainage deck and is able to drain considerable amounts of water at substantially zero headloss across the medium. Overall the diaphragm press operates like the recessed plate press, typically up to pressures between 690-1,040 kPa (100-150 psi).

Table 7-11. Performance Data for Solid Bowl Centrifuges

Plant	Sludge Type	Sludge Mixture P:S:C	Sludge Feed Rate	Sludge Feed Rate	Avg. Feed Solids Conc.	Avg. Cake Solids Conc.	Solids Recovery	Dewatering Chemicals		Chemical Cost	
		percent	gpm/unit	lb TS/d	%TSS	%TSS	percent	lb/ton DS	Type	$/lb	$/ton DS
Vidor, TX	Oxid. Ditch	100:0	40-60	500	1.0-4.0	17	98.9	26.7	Percol 767	1.65	44.06
Pinhole, CA	D (P + WAS)	50:50	45-65	3,600	1.5-2.0	14	95.0	10.0	Polymer	2.45	24.50
SSF, CA	D (P + WAS)	50:50	140-160	24,000	2.2-3.2	13[4]	90.0	13-15	Cat. Polymer	2.30	32.20
Port Huron, MI	R (P + WAS)	40:58.5:1.5[2]	35-61	11,122	2.2-6.9	20	94.0	9.0	Polymer	2.65	23.77
Oakland, CA	D (P + WAS)	40:60	150	140,000	2.0	20	85.0	18.0	Cyanamid E1125	0.98	17.64
San Rafael, CA	D (P + WAS)	50:50[2]	60-70	10,000	2.0-3.5	22	93.0	10.0 5.0	Allied Coll. $FeCl_3$	1.11 0.12	11.00 0.60
Valdese, NC	R (P + WAS)	45:55	40-75	1,495	3.0-5.0	18	90.0	16.0	Polymer	2.15	34.40
San Francisco, CA	D (P + WAS)	66:34[2]	125-250	145,000	1.0-3.0	18	92.0	8.9 50.0	Percol 757 $FeCl_3$	2.25 0.12	20.03 6.00
Detroit, MI	WAS	0:100	15	-[3]	1.5-3.5	15	89.0	80.0	Calloway 4450 Emul.	0.85	68.00
Blue Plains, DC	D (WAS)	40:60 (DP:RWAS)	150-200	142,809	5.3-6.8	17	98.4	10.1	Percol 757	1.71	17.27
Petaluma, CA	D (P + WAS)	80:20 (TF)	40	7,000	3.5-4.0	21	96.0	9.0	Percol 757	2.25	20.25
Denver, CO	D (P + WAS)	40:60	900	248,000	2.3-2.5	19	90.0	20.0	Percol 752	1.10	22.00

R = Raw; D = Digested; WAS = Waste Activated Sludge; P = Primary; S = Secondary; C = Chemical; TF = Trickling Filters.

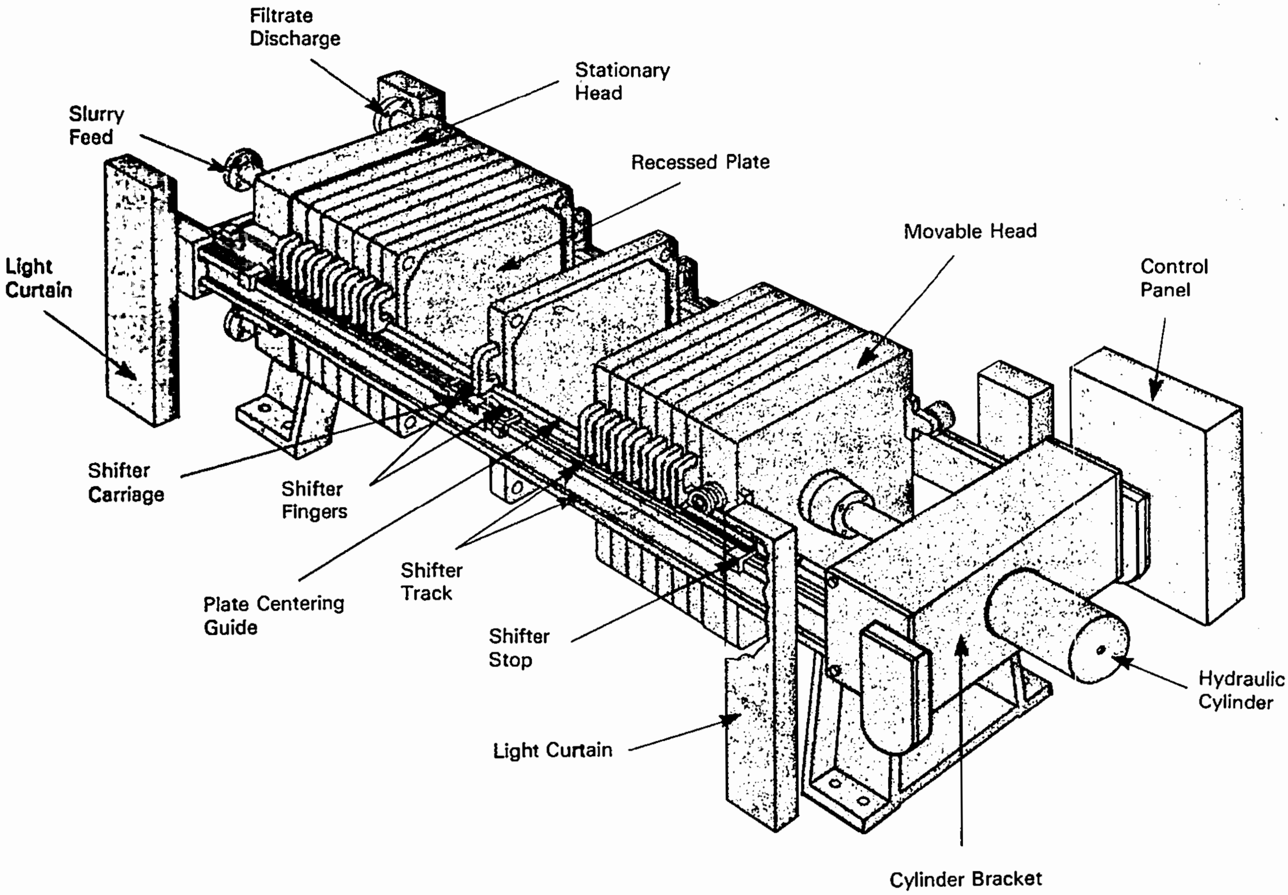
[1] 1986/1987 data.

[2] Unknown quantity of chemical P removal sludge in secondary sludge.

[3] Centrifuge only used when wasting rate is excessive and continued blending of primary and secondary sludges would overload belt filter presses.

[4] Sludge composted with rice hulls requiring wetter cake.

Figure 7-16. Fixed-volume recessed plate filter press (courtesy Eimco Process Equipment Co.).



The release of water at low pressures helps maintain the integrity of the floc. After water release appears complete following the initial filling period, pumping is stopped and the diaphragm cycle is initiated. The diaphragm pressure is applied, using either air or water on the reverse side of the diaphragm, and pressures up to 1,380-1,730 kPa (200-250 psi) are applied to the sludge for additional dewatering. In addition, the confined expression operation, which follows when the diaphragm pressure is applied effectively, releases substantial additional quantities of water (Cake solids will increase 5-8 percent).

A most significant aspect of the diaphragm press is that its construction and mode of operation allow the use of organic polymers as an alternative to ferric salts and lime conditioning techniques. Although there still is the same tendency to squeeze sludge into the media itself, the tendency is reduced by the elimination of substantial quantities of water prior to the start of the squeezing operation.

It should be noted that there has been an evolution in the diaphragm press to a simpler design sometimes known as a diaphragm plate press. In this design, both the cloth and the diaphragm are built into the plate. There fewer moving parts, longer cloth life, and much lower O&M costs.

Based on typical filtration operations, it can be expected that 70-85 percent of the water will be removed during the low pressure portion of the cycle of the recessed plate and diaphragm press. Similar performance can be obtained from a fixed-volume recessed plate press by stepping the pressure at two or three intermediate levels. The diaphragm press, however, usually produces a drier cake than that obtained from the fixed-volume recessed plate. Also, there is a substantially greater uniformity of solids concentration in the cake produced with a diaphragm press. With the low solids feed material continually being supplied to the recessed device, a very low solids cake fraction is produced near the feed point. This problem, of course, is not present in the diaphragm press because the pumping cycle is only the first part of the overall cycle and the diaphragm tends to remove water uniformly. Also, the cycle time for a given cake solids concentration is generally less in the diaphragm press.

There appears to be a less frequent need for precoating the diaphragm press than is usually encountered with the fixed-volume device. The implication is that the diaphragm press improves the dischargeability of the cake due to the higher cake solids content. Another advantage of the diaphragm press is that the sludge only needs to be pumped in at pressures up to, but rarely exceeding, 865-900 kPa (125-130 psi). The higher pressures during the diaphragm cycle may be supplied by clean water pumps or air pumps, thereby reducing the overall maintenance cost associated with high pressure delivery devices.

While polymers are uniquely successful in conditioning pure waste activated sludge and mixed sludges for dewatering in the diaphragm press, it would be misleading to say that most of the polymer conditioning success has occurred with these units. Actually, with mixed primary and secondary sludges, pure polymer conditioning has been most successful in low pressure presses operating at 520-1,040 kPa (75-150 psi), typically using one or two steps to achieve the ultimate pressure. However the low pressure recessed plate unit does not provide for the final high pressure water removal that the diaphragm press does and this could be the key to better cake discharge from the cloth. The operating sequences for fixed-volume recessed filter and diaphragm presses supplied by different manufacturers are shown in Figures 7-17 and 7-18, respectively.

7.4.3 Basis for System Design

This section provides an understanding of the following important properties:

- Cake solids concentration
- Throughput rate
- The recovery fraction, or the fraction of those solids delivered to the machine that exit the machine as cake and are not recycled to some other portion of the facility.

The cake solids concentration achievable with a particular sludge will regulate the cost of downstream operations and often determine the need for additional upstream operations such as thickening. There is a relationship between the cake solids concentration and the throughput in that, with filter presses, higher solids are almost always achievable. This is true for any given operating circumstance, if one is willing to increase the cycle time and, therefore, decrease the rate of throughput. The designer's challenge in this regard is to maximize throughput and solids concentration consistent with specific operating conditions.

The third critical design parameter is solids recovery. Systems that do not recover a substantial quantity of solids can experience an increased need for media washing and cause a buildup of fine solids in some process loop, especially one that goes to a thickener or perhaps to the wet end of the plant. Solids losses above 2-3 percent of feed solids are usually traced to torn media or sludge adhering to the media on discharge and washed off to be recycled. Sometimes this buildup of fines can lead to higher effluent suspended solid concentrations. In any event, recovery in excess of 95 percent is an important design objective of a system and is necessary to prevent both the excessive recycle of solids and the possible impact on some aspects of the wastewater treatment plant's operation. In this regard, filter presses generally are superior, with solids recovery typically greater than 98 percent.

7.4.4 Design Procedures

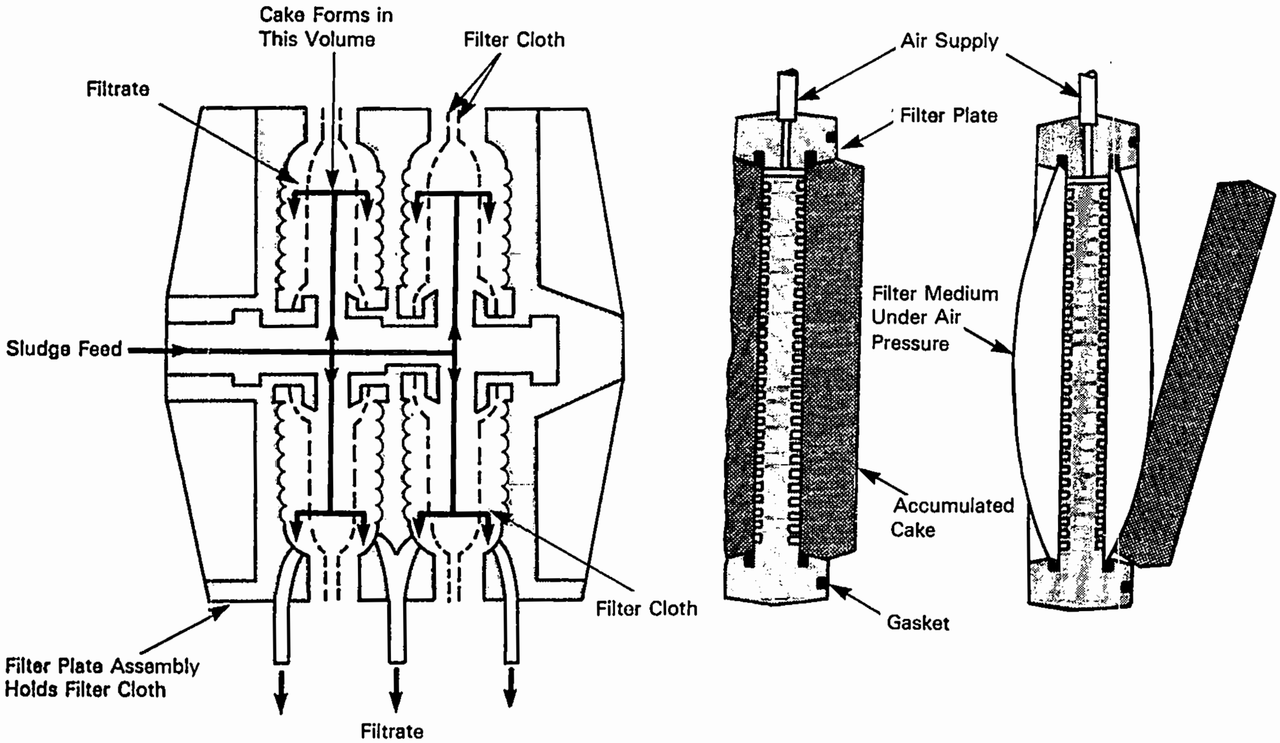
This section contains a review of the methods employed for predicting solids concentration, throughput, and recovery based on rather simple and straightforward laboratory tests. Pilot operations, if feasible, offer the best way of obtaining data on all three of the important design aspects. However, pilot operations are often not possible, in which case it becomes necessary to design from bench-scale information.

These procedures are based on one or both of two properties of sludge slurry systems and reflect the ease or difficulty in separating the water phase and the solid phase from each other. These properties are the Specific Resistance and the Capillary Suction Time (CST). These two properties have been defined and methods for their measurement are described in Sections 5.5.3.3 and 5.5.3.4 respectively. Specific Resistance and CST are used to develop design information on throughput and final solids concentration. Most of the relationships discussed below utilize the Specific Resistance test as the basic guide in estimating yield and cake solids. Yet, the CST is also a useful test, and this section of the manual contains several references to its use.

There are some significant dimensional considerations which must be discussed for a full understanding of Specific Resistance. Christensen (18) has summarized typical values of Specific Resistance for water and wastewater sludges and commented on the disparity in the use of units to describe Specific Resistance. He points out that sec^2/g probably is an incorrect unit assignment because of the manner in which these units were first employed. Gale (19) has pointed out that sec^2/g as a unit describing Specific Resistance was a result of improperly using g/cm2 for pressure difference across the filter cake. It has been suggested that meters per kilogram (m/kg) is a more satisfactory unit.

Christensen suggests the use of terameters per kilogram (Tm/kg) as the best possible unit. Christensen has also noted that proper conditioning, generally speaking, changes the Specific Resistance by a factor of 10^2 to 10^3. Raw wastewater sludges

Figure 7-17. Filling and cake discharge, fixed volume recessed plate filter press.



have Specific Resistance values of 10-100 Tm/kg. Adequately conditioned sludges have Specific Resistance values of about 1.0 Tm/Kg, and well conditioned sludges have Specific Resistance values on the order of 0.1 Tm/kg. The conversion factors to go from sec^2/g to cm/g and m/kg are 9.81×10^2 and 9.81×10^3, respectively.

The CST test provides a substantial amount of information about the ease in separating the water portion from the organic solids portion of sludge. For example, unconditioned waste activated sludge has a Capillary Suction Time of 100-200 seconds. For a filter press to function, dewater, and release the waste activated sludge cake, a Capillary Suction Time of 10 seconds or less is required.

The following series of relationships show the development of the significant equations that govern flow through porous media and, hence, filtration phenomena. In the 1800s, Poiseuelle described the velocity in a circular capillary tube as:

$$U = (d^2g/32\mu)(Pg/L) \qquad (7\text{-}25)$$

All in compatible units, where,

- U = linear velocity
- d = diameter of capillary
- P = pressure differential
- μ = viscosity of the liquid
- L = length of capillary
- g = gravitational constant

D'Arcy also in mid-1800s, showed that $(d^2g/32\mu)$ is a constant by noting that the flow through sand beds may be described by $U = (K_1) \times (P/\mu) \times (L)$, the other symbols are defined as before.

These equations were modified by Kozeny (20), who introduced porosity and specific surface in the equation:

$$U = \left[\frac{\varepsilon^3}{(1-\varepsilon)^2}\right]\left[\frac{1}{K\mu S_o^2}\right]\left[\frac{Pg}{L}\right] \qquad (7-26)$$

Again, in consistent values, where,

- ε = the porosity
- S_o = the Specific Surface
- K = a constant equal to 5

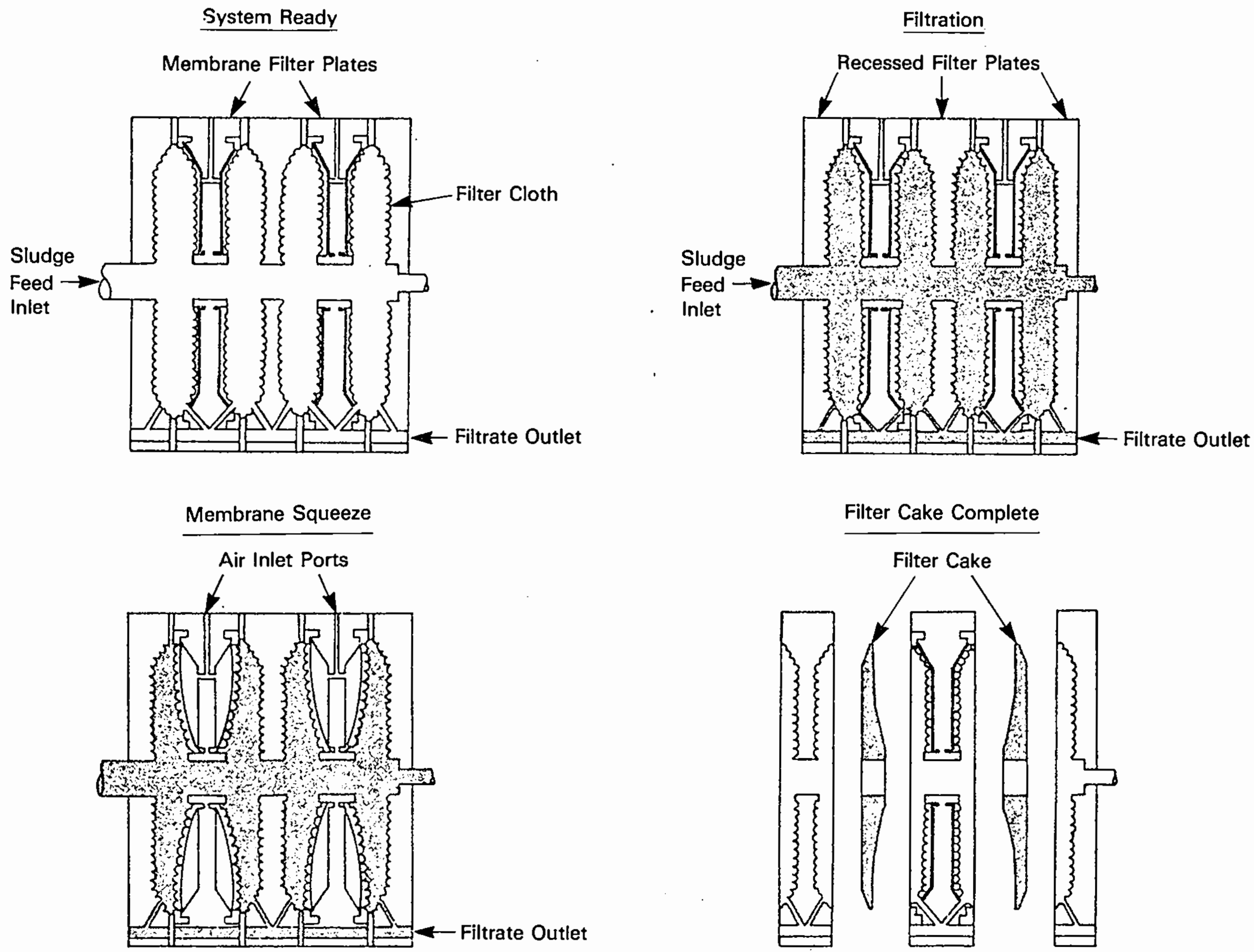
All other symbols are as indicated earlier.

Hence, with filtration:

$$U = (1/A)(dv/d\theta) \qquad (7\text{-}27)$$

and

Figure 7-18. Filling and cake discharge, diaphragm press.



$$U = K_1 \ (P/\mu L)$$

where,

K_1 = $(g\ \varepsilon^3)/[5\ (S_o)^2\ (1 - \varepsilon)^2]$
$dv/d\theta$ = rate of flow of liquid across cake

This form leads to the conventional expression:

$$dv/d\theta = PA/\mu RL$$

where,

L = cake thickness
R = specific resistance, sec^2/g
A = area of cake

but,

LA = cake volume also
v V = cake volume
v = volume of solids deposited per unit of filtrate

hence,

LA = v V
L = v V/A

and then by substitution,

$$dv/d\theta = PA^2/\mu RvV \qquad (7\text{-}29)$$

Carman (21), noting that R must include all resistance, developed an equation with two terms - one for the cake and one for the media:

$$dv/d\theta = PA^2/[\mu(rvV + R_m\ A)] \qquad (7\text{-}30)$$

where,

r = Specific Resistance of cake
v = cc of cake deposited by 1 cc of filtrate
R_m = initial resistance of 1 cm^2 of filtering surface

For compressible cake, vV becomes Vc where c is the weight of dry solids per unit volume in the unfiltered slurry. The general equation then becomes:

$$dv/d\theta = PA^2/[\mu(rcV + R_m A)] \qquad (7\text{-}31)$$

Integrating the expressions and neglecting the resistance of the media, the time for filtration becomes:

$$\theta = \mu rcV^2/2PA^2 \qquad (7\text{-}32)$$

If θ/V is plotted against V, a straight line is obtained whose V slope (b) is:

$$b = \mu rc/2PA^2 \qquad (7\text{-}33)$$

Therefore, the Specific Resistance may be calculated from Buchner funnel test data as described in Section 5.5.3.3, where,

$$r = 2bPA^2/\mu c \qquad (7\text{-}34)$$

Specific Resistance has been used in calculating yields from pressure filters. Coackley (22) reported a procedure in 1957. Mininni, Spinosa, and Misiti (23) have presented a procedure for predicting the filtrate flow rate and cake concentrations for fixed-volume pressure filter filtration. These workers observed that ϕ, the filtrate flow rate or flux, after the initial period of drainage or while the cake is being formed, is described by the expression:

$$\phi = at^b \qquad (7\text{-}35)$$

where t is time, and a and b are coefficients which can be determined if the Specific Resistance, initial solids concentration, filtrate viscosity, and maximum operating pressures are known. The final cake concentration then can be calculated by making a material balance, assuming that the dry solids density, the filtering time, the conditioner dosage, the slurry concentration, filter press chamber volume, and filtration surface areas are all known. The reported agreement between predicted and actual values is excellent.

Wilhelm (24) obtained the following expression from classical filtration theory:

$$\text{Log } \theta = \log [KA^2(S_c/c)^2] + \log [(\ell \rho_f)^2]$$

where,

- θ = filtration time, minutes
- A = filtration area, cm^2
- S_c = cake solids concentration by weight fraction
- c = feed solids concentration, gm/cm^3
- ℓ = cake thickness, cm
- ρ_f = feed density, gm/cm^3

Wilhelm's procedure provides excellent correlation when the cycle time is plotted against his correlating factor:

$$KA^2 (S_c)^2/c^2$$

To obtain good replication on different runs with the same sludge requires the solution of similar ranges of pressure to obtain "K" values.

The role of the filter media and the relationship between the character of the material being filtered and the media has been described in a study by Christensen and Sipe (25). They developed the following equation which, like the Carman equations, separates the resistance associated with the cake itself from that associated with the medium. The equation is:

$$\frac{t}{V} = \frac{\mu Rc}{2P_tA^2}\left[V + \frac{\mu R_m}{P_tA}\right] \qquad (7-36)$$

where,

- t = time
- V = filtrate volume
- μ = absolute viscosity
- R = Specific Resistance (in the case of m/kg)
- c = mass of cake deposited per unit volume of filtrate
- P_t = total pressure drop across cake and medium
- A = filtration area
- R_m = resistance of the medium

If the medium resistance is negligible, the preceding equation can be rearranged to a form similar to Carman's equations:

$$\log t = (\mu Rc/2P_cA^2) + 2 \log v \qquad (7\text{-}37)$$

where,

P_c = Pressure drop across cake

The authors suggest that the rearrangement of the equation offers the second way to plot filtration data, i.e., log t versus log v. When this is done, according to the equation, the data should plot as a straight line with the slope of two. The intercept at a convenient point, such as volume = 1, can be used to calculate the Specific Resistance, since all of the other factors in the first term of the equation are known. The authors also point to many significant deviations when the slope is equal to two; one is that the equation can be written as $t = KV^n$, where n varies. In the t/V vs. V plot (which will be linear only if n is equal to two when the data is approximated by a straight line), the intercept of that line will be negative when n is greater than two and positive when n is less than two. The t/V intercept is proportional to the medium resistance.

Constructing a log t versus log V plot of the data from a filtration experiment with significant medium resistance is equivalent to satisfying the equation: log $t = \log (K_1V^2 + K_2V)$ where K_1 represents the constant items in the initial term of the first of equation of Christensen and Sipe and where K_2 represents the constants in the second term.

Noting a consistent deviation from theoretical practice, Notebaert et al. (26) have proposed a modification of the standard cake filtration model to account for the deviations. The significant findings of Notebaert et al. were summarized by Christensen and Sipe (25) as follows:

- The assumption that the medium does not become fouled or clogged is not realistic.

- If the particles in the sludge are of the same order as the pores in the filter medium, the medium will clog. If the particles are a great deal larger than the pores in the medium, clogging will still occur but it will occur over a much longer period.

- If the medium clogs, resistance will be high at the start of the filtration cycle while the medium is clogged, but will increase slowly afterwards. Therefore, the average Specific Resistance will be decreasing throughout filtration. If the cake becomes clogged, the clogging will continue throughout filtration with a continuous increase in the average Specific Resistance.

- The slope of the log t versus log V plot is indicative of the physical processes described. If the medium is clogging, the slope will be less than two. If the cake is clogging, the slope will be greater than two.

Selection of the optimum filter media, based on the manufacturer's specification characteristics of the media (which will include data such as air flow, the weave, the fabric, etc.) is not yet possible. However, the Metropolitan Waste Control Commission of the Twin Cities (27) has carried out a detailed and comprehensive study on pressure filtration. When wastewater, without solids, comes in contact with the filter medium, the media resistance will increase. This is probably due to bacterial growth, since the presence of chlorine decreases the rate at which resistance increases. Pressure and the impact of pressure on the fibers themselves increases the extent and rate of blinding. The workers observed that polypropylene and nylon are the two most commonly used materials for filter cloth. The authors, in their literature review, pointed to Purchas' work. He tried to relate filtrate clarity, resistance to flow, cake solids, ease of discharge, cloth life, and tendency to blind to media characteristics. They also noted that criteria set forth by Warring might be the best and most reliable guide for establishing a good model. These criteria are:

- How small a particle can the media retain?

- What is its resistance to flow?

- What is the relationship between buildup of particulates in the medium to the rate of flow?

Finally, these workers concluded from their own studies that the resistance of the filter cloth increased markedly with use. Periodic washing with water or with acid reduced media resistance. The effect of media resistance on press operation was found to be very significant. Cake solids decreased from 60 percent to 33 percent on a full-scale press due to increased media resistance. Filtration rates, filtration yield, mass of dry solids deposited, and cake percent solids all decreased because of increases in media resistance. Any model of a pressure filtration process must include a term for media resistance.

7.4.5 Support Equipment and Processes

Sludge must maintain some structural integrity during the pressing period, since a massive structure will prevent the movement of water through the filter cake to the discharge or filtrate side. To this end, one of the essential parts of a filter press system is the sludge conditioning subsystem. For recessed volume filter presses, the most common conditioning technique for digested and waste activated sludges - and probably the most common for primary sludge - is the addition of iron salts and lime. The average quantity required is on the order of 5 percent ferric chloride and 20 percent lime, though values as low as 3 percent ferric chloride and 10 percent lime and as high as 10 percent ferric chloride and 40 percent lime have been reported and are sometimes required.

In general, the ferric chloride requirement is a function, at least in part, of the sludge alkalinity. The role of lime in sludge conditioning has been discussed extensively by Webb (28) and Sontheimer (29), but is still somewhat unclear. Lime's solubility above pH 11 or 12 is only on the order of one gram per liter. As a result, much of the lime must exist as partially hydrated calcium hydroxide, which probably acts structurally to provide channels for water to move through to the filtrate side.

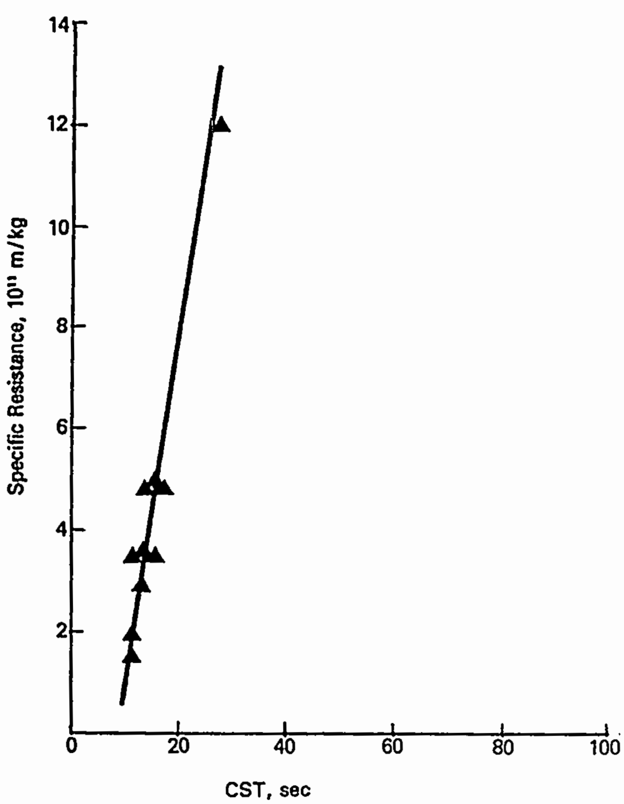
In studies with four iron salt conditioners and lime, Christensen and Stule (30) obtained both CST data and Specific Resistance data. The results are shown in Table 7-12 together with a correlation between these two for the particular sludge under study. This data is presented in Figure 7-19 and shows a remarkably good correlation. Briefly, the ferric conditioners performed the best and the chloride form was more effective than sulfate ore which appeared

Table 7-12. **Comparison of Iron Conditioners With and Without Lime (14)**

Total Sludge Solids[1]	Iron Conditioner	Iron Dose	CST After Iron Addition	Lime Dose	Specific Resistance After Iron and Lime Addition
percent		percent	sec	percent CaO	10^{11} m/kg
5.5	$FeSO_4 \bullet 7H_2O$	1.72	208	15	14.0
5.5	$FeCl_2 \bullet 4H_2O$	1.72	157	15	7.9
5.5	$Fe_2(SO_4)_3 \bullet 6H_2O$	1.72	41	15	5.0
5.5	$FeCl_3 \bullet 6H_2O$	1.71	26	15	2.6
5.5	$FeSO_4 \bullet 7H_2O$	3.44	180	30	6.0
5.5	$FeCl_2 \bullet 4H_2O$	3.44	139	30	2.9
5.5	$Fe_2(SO_4)_3 \bullet 6H_2O$	3.44	27	30	2.3
5.5	$FeCl_3 \bullet 6H_2O$	3.44	19	30	1.2
7.0	$FeSO_4 \bullet 7H_2O$	3.44	480	20	11.0
7.0	$FeCl_2 \bullet 4H_2O$	3.44	[2]	20	5.6
7.0	$Fe_2(SO_4)_3 \bullet 6H_2O$	3.44	117	20	5.3
7.0	$FeCl_3 \bullet 6H_2O$	3.44	58	20	1.8

[1] The activated sludge employed was a raw mixed sludge approximately 50% primary and 50% waste activated on a dry solids basis.
[2] No result because of a lab accident.

Figure 7-19. **Specific Resistance vs. Capillary Suction Time.**



to produce a slightly poorer floc and a more poorly conditioned cake than did the chloride.

The role of calcium was also studied extensively, and it was concluded that calcium is involved in a chemical link with the iron floc. However, calcium chloride was used and the pH was raised with sodium hydroxide to obtain a somewhat synthetic situation. Extensive reports exist which relate the necessity to clean calcium hydroxide scale off both media and plates, indicating that considerable quantities of calcium hydroxide exist when the sludge is conditioned with lime. Two other important observations are that aging the sludge can as much as double the Specific Resistance in one hour. The actual impact of aging depends on the flocculants used. Also, it is clear that the pH has a substantial effect. The Specific Resistance dropped from 1.6 x 10^{12} m/kg at a pH value of around 11.3 to about 0.2 x 10^{12} m/kg at a pH of 12.45. This clearly is a pH related phenomenon and not tied into any physical property of calcium because no precipitation of calcium hydroxide was observed.

The use of polymers for sludge conditioning is expanding. Polymers can produce very nearly the same cake solids and do not result in a 15-30 percent increase in cake weight and volume. Dosing procedure, flocculation requirements, and filter press pressure-time relationships necessary to optimize polymer dosage and cake release are site specific. About 75-80 percent of the conversion to polymer trials appear successful. Polymer costs are 30-70 percent of ferric and lime costs.

7.4.6 Operational Factors and Performance Characteristics

This section deals with the machine and process variables that affect the efficiency of filtration when dewatering with a conventional recessed plate filter press or a diaphragm filter press. The first part is devoted to machine variables, which are developed or derived from the unique and special characteristics of the machine in use. The second part discusses those variables that arise from the process or the unique characteristics of the sludge to be filtered.

In a conventional filter press, the operator controls the following variables:

- Pressure of the feed sludge
- The rate at which the pressure is applied and the pacing of flow to the filter press
- The overall filtration time, including such variables as the time at each pressure level in multiple pressure level operations
- The use of precoat or body feed and the amount of material used
- Conditioning chemicals
  - Type
  - Dosage
  - Location
  - Mixing efficiency
  - Flocculation efficiency
- Cloth washing frequency
- The nature of the filter media used.

A similar set of machine variables exists for the diaphragm filter press. They are:

- Pressure of the feed sludge and the rate at which feed sludge is added to the machine.
- Filtration time
- Diaphragm pressure
- Diaphragm squeezing time
- Rate at which the diaphragm pressure is increased
- Conditioning chemicals
  - Type
  - Dosage
  - Point of addition
  - Mixing efficiency
  - Flocculation efficiency
- Filter media used
- Cloth washing frequency.

Changes in these parameters are predictable up to a point, and mechanisms exist to evaluate the effect of varying each one for optimizing the system.

Precoat generally does not need to be used when inorganic conditioning chemicals, particularly ferric chloride and lime, are used. Heavy doses of organic polyelectrolyte may also preclude the use of body feed or precoat. Precoat is normally used in cases where the particle size is extremely small or considerable variability in filterabilty and substantial loss of fine solids to and through the filter media are anticipated. A final decision about the need for precoating may require lab or field experimentation with the specific sludge.

When substantial quantities of lime are used, cloth washing may require both an acid and a water wash. Therefore, a medium is needed that is resistant to both acid and alkaline environments. In those instances where polyelectrolytes are used, the washing operation normally is accomplished with only clean water since the sludge imbedded in the media is backflushed to the waste.

Few process variables, as opposed to machine variables, are likely to be controllable by the operator. Process variables include:

- The type of sludge to be dewatered. Raw or digested primary sludge, waste activated sludge, trickling filter sludge, RBC sludge, or mixtures thereof have varying effects on the dewatering process.
- The age or the freshness of the sludge. Conditioning, particularly conditioning with polyelectrolytes, is much more dependable and reproducible when the sludge is fresh. The Specific Resistance increases with time. Therefore, it is desirable to dewater the sludge in as fresh a condition as possible.
- Prior chemical conditioning. Prior chemical conditioning tends to confound the use of chemical conditioning at the dewatering device. This is particularly true when polymers are used and if polymers have already been used somewhere upstream from the dewatering system. If this condition exists, the best remedy is to use a small quantity of the polymer of the opposite or neutral charge, followed by the normal dose of the polymer usually employed. Establishing charge reversal with the polymer of the opposite charge eliminates the confounding effects of the old, partially degraded polymer on the sludge surface.
- The solids concentration achievable in the final clarifier or in subsequent thickening operations. Generally speaking, it is desirable to send to the dewatering device a feed sludge with the highest possible solids content.
- Solids capture. If the cloth is unbroken and cake cleanly discharged, suspended solids recovery is

about 99 percent. When the cloth is washed, the effluent solids are somewhat higher.

- Cake concentration. The cake concentration must be sufficiently high to readily discharge from the cloth. Variables affecting cake concentration have been reviewed earlier.

- Throughput rate. The throughput will be dependent on the water release characteristics of sludge, type and amount of chemicals, and the desired minimum cake solids.

- Conditions under which the sludge was produced. The filterability of sludge, particularly waste activated sludge, is strongly dependent on the conditions under which the sludge was produced. This consideration probably applies to those municipal wastewater treatment plants receiving substantial quantities of high carbohydrate industrial wastes that may produce, on occasion, a nitrogen deficient situation in the activated sludge portion of the plant. However, nitrogen deficient activated sludge has a considerably higher Specific Resistance when untreated than activated sludge grown under nitrogen enriched conditions. In addition, the final Specific Resistance after chemical conditioning is not as good as that achieved with activated sludge grown under excess nitrogen conditions. The conditioned Specific Resistance of the nitrogen- poor sludge generally runs two to three times that of the activated sludge grown under high nitrogen conditions when properly chemically treated (31).

7.4.7 Survey of Filter Presses

In order to provide detailed information on the operating experience of filter presses, a survey was made of 50 municipal wastewater treatment plants by Terraqua Corp. of Hunt Valley, MD (32) for the City of Baltimore in 1984. An effort was made to update the information to April 1987. The plants ranged in size from 0.66 to 265 m^3/min (0.25 mgd to 100 mgd).

Table 7-13 summarizes general information on the filter press installations. Of the 50 plants, 42 were dewatering anaerobically digested or raw sludges: 21 of each sludge type. One plant was elutriating anaerobically digested sludge, five were processing aerobically digested sludge, one was thermally conditioning anaerobically digested sludge, and one was dewatering an alum sludge from a tertiary treatment process.

The majority of plants, 41 of them, had their presses in operation at the time of the study. Of the remaining plants, four had taken their presses out of service (one temporarily, three permanently), two planned to take them out of service, two had presses under start-up, and one had not yet completed installation. Most of the plants were landfilling or incinerating their filter cake.

Table 7-13. Summary Data - General Information

	Number of Plants
Type of Sludge Processed	
Anaerobically digested	21
Anaerobically digested/elutriated	1
Raw	21
Aerobically digested	5
Other (thermally conditioned and chemical)	2
Total	50
Operating Status of Presses	
In service	41
Abandoned use of press	3
Out of service temporarily	1
Planned to be taken out of service	2
Under startup	2
Under construction	1
Total	50
Filter Cake Disposal Method[1]	
Landfill	24
Incinerate	11
Land Apply	5
Incinerate/Landfill	3
Landfill/Land Apply	2
Compost/Land Apply	1
Total	46
Press Manufacturers	
Passavant	14
Edwards & Jones	12
Eimco(Shriver)	11
Sperry	4
Netzsch	3
Hoesch	3
Envirex (NGK)	1
Clow	1
Ingersoll-Rand (Lasta)	1
Total	50
Longest in continuous operation:	
Statesville, NC - since 1974 - Passavant	
Plate Material[2]	
Cast Iron	33
Polypropylene	14
Not reported	4
Total	51
Operating Experiences[3]	
Positive	29
Negative	12
Mixed	8
Total	49

[1] Excluding plants under construction or which have abandoned use of press.
[2] One plant has both plate types.
[3] Excluding one plant under construction.

Three manufacturers of recessed chamber presses dominated the installations: Passavant, Edwards and Jones, and Eimco (Shriver press). Other recessed-chamber press manufacturers with municipal installations included Sperry, Netzsch, Hoesch, and Clow. Two diaphragm press manufacturers were surveyed - Envirex (NGK press) and Ingersoll-Rand (Lasta press). The manufacturer with the longest continuous operating history was Passavant, which has an installation at the Statesville, NC plant operating since 1974. The majority of installations were relatively recent, since about 1980. Most of the presses used cast iron plates (33 plants), but polypropylene plates were used by some manufacturers for new presses; polypropylene had been used as a replacement for cast iron plates by some plants.

Based on the telephone conversations and on the written comments received from a mail survey, a rating of positive, negative, or mixed was given to the attitude of operations personnel toward operating and maintaining their filter press system. In general, a negative rating was assigned to plants which have taken their press system out of service due to excessive operating and/or maintenance costs, which reported serious maintenance problems, or which had a very negative reaction toward the installation. A mixed rating was given to plants which had less serious operating problems and where the operator's attitude was more positive than negative. A positive rating was given where the personnel were generally satisfied or, in some cases, enthusiastic about the press, although some problems may have been reported. The experience of the majority of the plants was rated as positive (29 of 49 plants, excluding one plant under construction). Twelve had negative ratings; of these, four plants have ceased using the press. Eight plants had mixed operating experience.

Table 7-14 presents a summary of operating comments itemized by type of problem and grouped by the overall rating of operating experience. A large number of plants (28) reported no significant problems with the press installation. It was commonly reported even by plants with both positive and negative overall attitudes toward the press system that costs were high for operating labor and for chemical conditioning, and that well trained, motivated operators and mechanics were a necessity. Plants with positive attitudes seemed to be able to overcome this difficulty by good training and supervision, while plants where the attitude was negative seemed, in contrast, to be overcome by it.

Table 7-15 presents a summary of data on conditioning chemical dosages and filter cake quality. Most of the plants, 29 of them, were conditioning with ferric chloride and lime only, six were using a precoat of ash or diatomaceous earth with ferric and lime conditioning, and six were conditioning with polymer alone. The less frequent conditioning methods included ferric chloride/lime with ash and polymer/precoat with and without ash. One plant was using lime alone, but was dewatering an alum sludge from a tertiary phosphorus removal system.

Reported filter cake solids content averaged 37 percent for the plants using ferric/lime conditioning and only slightly less at 34 percent for plants using polymer alone. Plants using a precoat reported one of the highest solids contents, 42 percent with ferric and lime, but this method had the disadvantage of maximizing performance at the expense of additional inert material in the cake. The two plants using ash as a conditioning material reported 45 percent with polymer and a precoat and 32 percent with ferric and lime. However, the amount of ash added, from 63 percent to 100 percent, adds significantly to the amount of filter cake to be disposed. Two plants were using polymer with a precoat and reported an average cake solids of 33 percent. Table 7-16 presents a summary of filter press performance on anaerobically digested and raw sludges for the three most common conditioning methods (ferric/lime, polymer, and ferric/lime/precoat). Surprisingly, in each case, the best results were obtained on digested rather than raw sludge: 38 percent vs. 36 percent for ferric/lime, 35 percent vs. 31 percent for polymer, and 44 percent vs. 32 percent for ferric/lime/precoat. This data should be interpreted carefully, however because numerous other factors can influence press performance. Such factors include operating pressure, cloth condition, feed solids percent, primary/secondary sludge ratio, chemical dosage, and press cycle time. An example is Watertown, NY, which reported a 37 to 44 percent solids cake with polymer-only conditioning, but required 22.5 kg/Mg (45 lb/ton) of polymer and a very long (4-hr) cycle time.

Chemical dosages and cycle times are also listed in Table 7-16. As shown, chemical dosages for anaerobically digested sludges averaged (in order by conditioning method): (1) 7 percent ferric, 26 percent lime without precoat; (2) 18.5 kg/Mg (37 lb/ton) polymer; and (3) 9 percent ferric, 32 percent lime with precoat. Chemical dosages for raw sludges averaged: (1) 7 percent ferric, 23 percent lime without precoat; (2) 6 kg/Mg (12 lb/ton) polymer; and (3) 8 percent ferric, 20 percent lime with precoat. Average cycle times varied from about 1.5 to 2.5 hr.

7.4.8 General Equipment Selection Criteria

Ohara et al., (33) in writing about the Hyperion system, developed the following set of criteria, in order of priority, for selecting the most cost effective, functional, safe and environmentally sound system:

- Meets all environmental and legal requirements
- Has minimum energy, resource and economic requirements
- Minium suspended solids remain on the liquid side stream, whether it be concentrate, filtrate, or supernatant
- Provides capture of sludge solids
- Provides maximum cake solids (minimum percent moisture in the cake)
- Has maximum operational reliability, flexibility, and ease of use

Table 7-14. Summary of Operating Problems

Item	No. of Plants (by overall rating)			Total No. of Plants Reporting
	Positive	Negative	Mixed	
No significant problem	11	-	-	11
High maintenance costs or unspecified mechanical problems	1	5	2	8
High operating or conditioning costs	4	4	3	11
Well trained and/or motivated operators and mechanics needed	4	4	-	8
Sludge feed problems (line clogs, feed pumps)	2	1	3	6
Presselectrical or instrumentation programs	4	1	-	5
Excessive cloth wear or tears	1	1	4	6
Conditioning system problems (corrosion, line clogs, poor uniformity or conditioning)	1	2	2	5
Stayboss wear or failure	1	-	3	4
Plate suspension pin breakage	1	-	3	4
Difficult to obtain spare parts	2	2	-	4
Ammonia release problem	1	1	2	4
Plate shifting mechanism problem	2	1	-	3
Cake discharge system problems (conveyors, drip trays)	1	1	1	3
Plate cracking or breakage	-	1	1	2
Hydraulic power unit leakage	-	1	-	1
Poor cake solids	-	1	-	1
Press frame twisting	-	-	1	1
Plate coating wear	1	-	-	1
Rapid cloth blinding	-	1	-	1
Filtrate drain lime accumulation	-	-	1	1

Table 7-15. Filter Cake Solids - Average by Conditioning Method

	No. of Plants[1]	Percent Solids	
		Mean[2]	Std. Dev.
Ferric lime only	29	37	5.3
Ferric/lime/precoat	6	42	7.8
Lime only (alum sludge)	1	38	-
Ferric/lime/ash	1	32	-
Polymer only[3]	6	34	4.2
Polymer/precoat	2	33	4.2
Polymer/ash/precoat	1	45	-

[1] Exculding plants that have abandoned use of press or are under construction.
[2] Using midpoint data for plants reporting a range of values.
[3] Excuding thermally conditioned sludge.

- Requires minimum maintenance and downtime
- Has maximum flexibility to meet changing needs
- Can meet established construction schedules.

Table 7-16. Filter Cake Solids - Average by Conditioning Method and Sludge Type[1]

Conditioning Method and Sludge Type	No. Plants Reporting	Average Cake Solids	Ferric Only (%)			Lime Only (%)			Polymer Only (lb/ton)			Cycle Time[2] (minutes)		
		percent	Min	Max	Avg	Min	Max	Avg	Min	Max	Avg	Min	Max	Avg
Ferric/lime														
Anaerobic	11	38	3	15	7	12	43	26				52	168	100
Raw	14	36	1.5	12	7	15	30	23				70	330	155
Polymer														
Anaerobic	2	35							20	45	37	60	240	150[3]
Raw	3	31							7	18	12	50	105	82
Ferric/lime/precoat														
Anaerobic	4	44	7	10	9	16	40	32				90	105	99
Raw	1	32	-	-	8	-	-	20				-	-	120

[1] Using midpoint of data for plants reporting a range of values.
[2] Excuding diaphragm press installations.
[3] Only two plants reporting, wide variation.

7.5 Vacuum Filtration

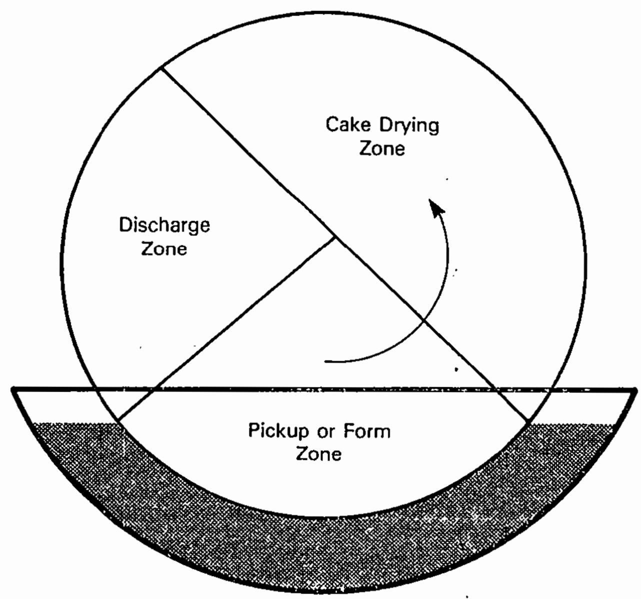
7.5.1 Introduction

The most common means of mechanically dewatering municipal wastewater sludge until the mid-1970s was vacuum filtration. Vacuum filters were patented in England in 1872 by William and James Hart. The first United States application of a vacuum filter in dewatering municipal wastewater treatment plant sludge was in the mid-1920s. Until the late 1950s, the drum or scraper-type rotary vacuum filter was the most common design of vacuum filter employed. The belt-type filter using stainless steel (SS) coils was introduced by Komline Sanderson in 1951. Since then and until the mid 1970s, the belt-type filter with natural or synthetic fiber cloth, woven SS mesh, or coil springs media has been the dominant means of mechanically dewatering sewage sludge.

A vacuum filter consists of a horizontal cylindrical drum which rotates while partially submerged in a vat of sludge. The filter drum is partitioned into several compartments or sections. Each compartment is connected to a rotary valve by a pipe. Bridge blocks in the valve divide the drum compartments into three zones, which are referred to as the cake formation zone, the cake drying zone, and the cake discharge zone.

The filter drum is submerged to about 20-35 percent of its depth in a vat of previously conditioned sludge; this submerged zone is the cake formation zone. Vacuum applied to this submerged zone causes filtrate to pass through the media and sludge particles to be retained on the media. As the drum rotates, each section is successively carried through the cake formation zone to the cake drying zone. This zone begins when the filter drum emerges from the sludge vat. The cake drying zone represents from 40 to 60 percent of the drum surface and ends at the point where the internal vacuum is shut off. At this point, the sludge cake and drum section enter the cake discharge zone, where sludge cake is removed from the media. Figure 7-20 illustrates the various operating zones encountered during a complete revolution of the drum.

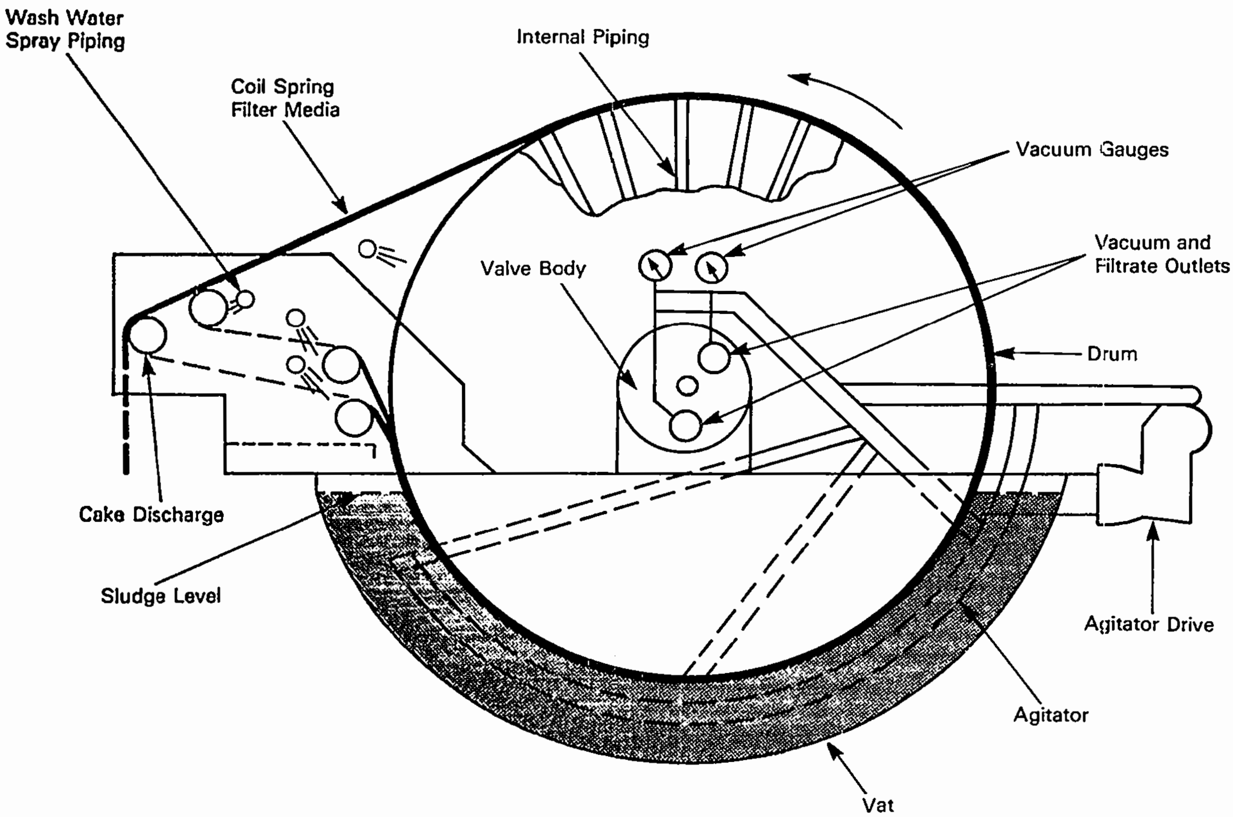
Figure 7-20. Operating zones in a rotary vacuum filter.



There are essentially two variations currently on the market: the drum filter and the belt filter. In the case of the drum filter, the media covers the drum and the sludge is removed by a roll discharge or a doctor blade. The belt filter may employ conventional media or coil media, but when conventional media are employed, the belt leaves the drum for discharge and is washed before recontacting the drum.

Figure 7-21 shows a cross sectional view of a coil spring, belt-type vacuum filter. This filter uses two

Figure 7-21. Cross-sectional view of a coil spring, belt type rotary vacuum filter.



layers of stainless steel coils arranged around the drum. After the cake drying or dewatering cycle, the two layers of springs leave the drum and are separated from each other. In this way, the cake is lifted off the lower layer of springs and can be discharged from the upper layer. Cake release from the coils is usually not a problem if the sludge is properly conditioned. After cake discharge, the coils are spray washed and returned to the drum just before the drum reenters the sludge vat.

The coil springs, which have 7 to 14 percent open area, act to support the initial solids deposit which in turn serves as the filtration medium. Because of the open area of the springs, it is important that the feed solids concentration be high; that is, it should contain sufficient fibrous material to prevent the loss of fine solids. Sludges with particles that are both extremely fine and resistant to flocculation dewater poorly on coil filters, and solids capture is low. A cloth medium is required when filtering unthickened sludge that is predominantly secondary solids.

Figure 7-22 shows a schematic cross section of a fiber cloth, belt-type rotary vacuum filter.

In this type of unit, the media leave the drum surface at the end of the drying zone and pass over a small-diameter discharge roll to facilitate cake discharge. Washing of the media occurs after discharge and before return to the drum for another cycle. This type of filter normally has a small-diameter curved bar between the point where the belt leaves the drum and the discharge roll. This bar aids in maintaining belt dimensional stability and ensures adequate cake discharge. Scraper blades, additional chemical conditioner, or the addition of fly ash are sometimes required to obtain cake release from the cloth media. This is particularly true at wastewater treatment plants that produce sludges that are greasy, sticky, and/or contain a large quantity of waste activated sludge. In general, cloth media made from staple fiber produces cleaner filtrate but has lower throughput than cloth media made from monofilament fiber.

7.5.3 Design Procedures

The best way of carrying out bench-scale studies involves the use of Capillary Suction Time, Specific Resistance, and Filter Leaf Tests. These test procedures are all described in detail in Chapter 5.

7.5.4 Support Equipment

Vacuum filters are normally supplied with auxiliary equipment including vacuum pump, vacuum filtrate receiver and pump, and sludge conditioning apparatus. Figure 7-23 shows a typical complete rotary vacuum filter system.

Figure 7-22. Cross-sectional view of a cloth, belt type rotary vacuum filter.

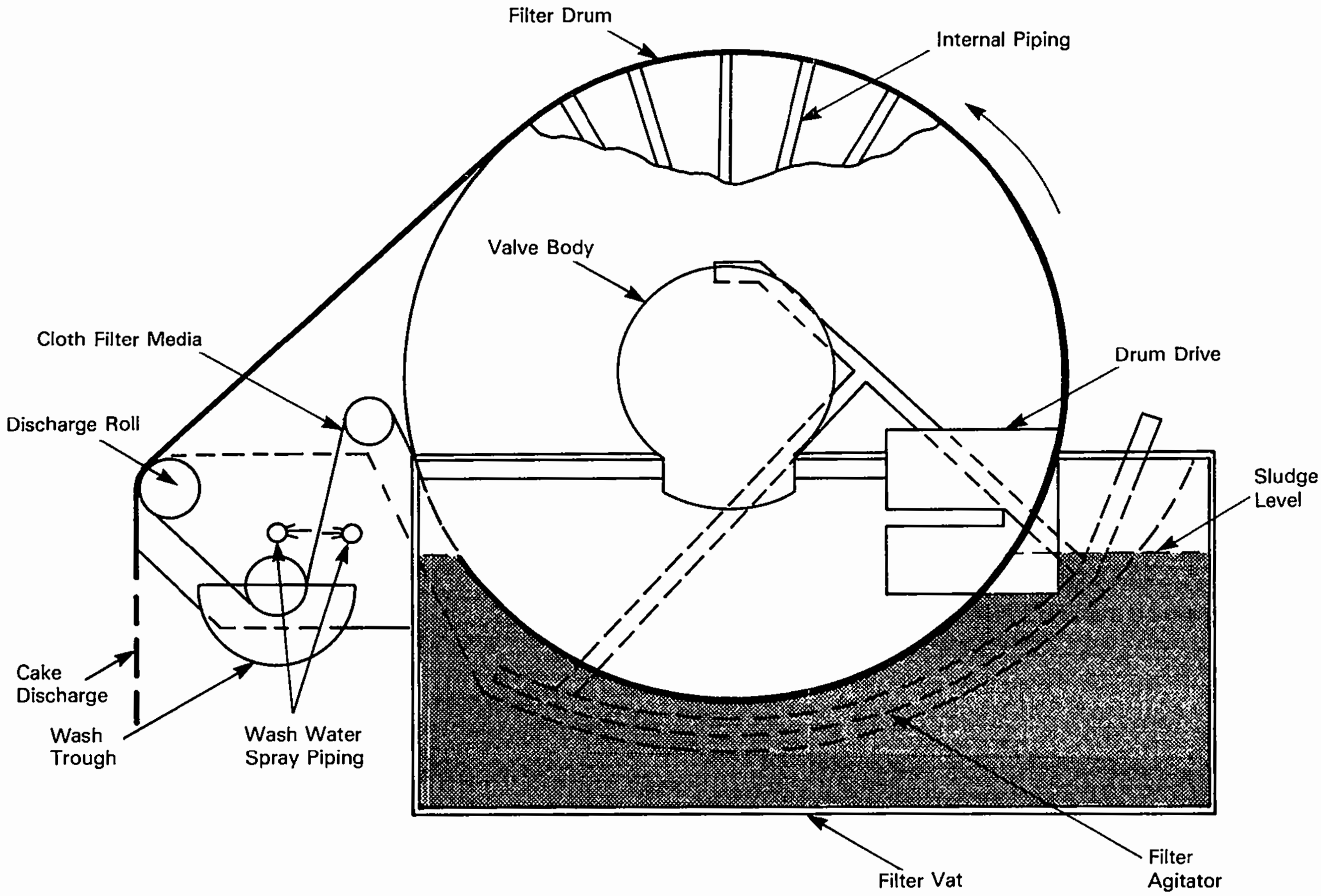
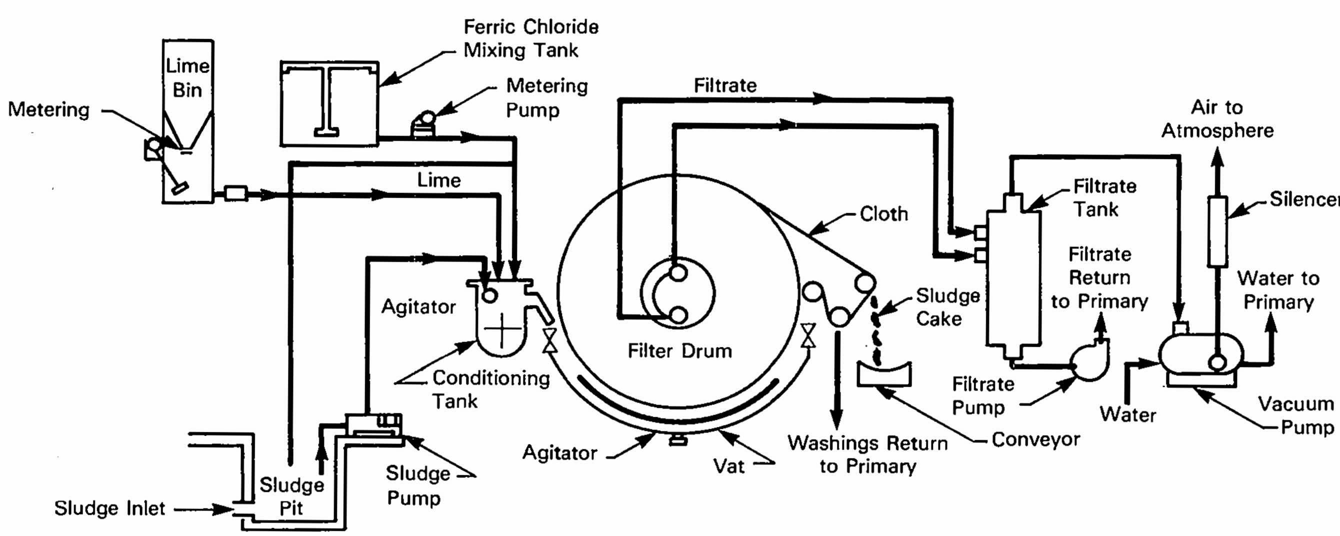


Figure 7-23. Rotary vacuum filter system.



Usually, one vacuum pump is provided for each vacuum filter, although some larger plants use fewer than one pump per filter with the pumps connecting to a common header. Until the 1960s, reciprocating-type dry vacuum pumps were generally specified, but since the early 1970s wet-type vacuum pumps have been universally used. The wet-type pumps are more easily maintained and provide sufficient vacuum. Wet-type pumps use seal water, and it is prudent to use potable water. If the water is hard and unstable, it may be necessary to prevent carbonate buildup on the seals through the use of a sequestering agent. The vacuum pump requirement is normally 0.7-1.0 m^3/min of air /m^2 of drum surface area at 33 kN/m^2 absolute pressure (1.4-2.0 cfm/ft^2 @ 5 psi). If the expected yield is greater than 20-40 kg/m^2 hr (5-10 lb/ft^2 hr) and extensive sludge cake cracking is expected, an air flow 2.0-2.5 times higher should be used.

Each vacuum filter must be supplied with a vacuum receiver located between the filter valve and the vacuum pump. The principal purpose of the receiver is to separate the air from the liquid. Each receiver can be equipped with a vacuum-limiting device to admit air flow if the design vacuum is exceeded (a condition that could cause the vacuum pump to overload). The receiver also functions as a reservoir for the filtrate pump section. The filtrate pump must be sized to carry away the water separated in the vacuum receiver, and it is normally sized to provide a capacity two to four times the design sludge feed rate to the filter.

The filtrate pump should be able to pump against a minimum total dynamic head of between 12-15 m (40-50 ft), which includes a minimum suction head of 7.5 m (25 ft). Centrifugal pumps are commonly used but can become air-bound unless they have a balanced or equalizing line connecting the high point of the receiver to the pump. Typically, non-clogging centrifugal pumps are used with coil filters because they permit a somewhat higher solids concentration in the filtrate. Self-priming centrifugal pumps are used most frequently, since they are relatively maintenance free. Check valves on the discharge side of the pumps are usually provided to minimize air leakage through the filtrate pump and receiver to the vacuum pump.

7.5.5 Operating Factors and Performance Characteristics

7.5.5.1 Machine Variables

The principle machine variables that impact on vacuum filter operation are as follows:

- Filter media used
- Quantity of wash water used
- Drum speed
- Vacuum level
- Conditioning chemicals - type and dosage
- Drum submergence
- Vat agitation.

Establishment of the drum speed, optimum vacuum level, conditioning methods, drum submergence and optimum media selection can all be accomplished on bench scale. The drum speed establishes the cycle time and the submergence sets the form time and drying time. The media selection is normally made at the time of equipment start-up by the equipment supplier. The trend over the past few years has been to select a monofilament fabric, since they seem the most resistant to blinding and have a reasonably long life.

A change in conditioning procedures. sludge mixture, or sludge holding time (time held before conditioning and dewatering) impact on the efficiency of a given medium.

7.5.5.2 Other Process Variables

With belt filters, wash water at a pressure of at least 480 kN/m^2 (70 psi) must be available. Throughput is usually estimated from data gathered with clean media. It is generally observed that where there is insufficient cloth washing, increasing the amount of wash water will increase the machine throughput and will help to increase cake dryness. Vat agitation is necessary for proper cake formation, but over-agitation will result in breaking up the sludge floc, poor solids capture, and lower feed rates. The addition of scraper blades, use of excess chemical conditioner, or addition of fly ash are sometimes required to obtain cake release from cloth media vacuum filters.

The feed solids concentration has a critical effect on the filter's production rate and the final solids concentration. The higher the suspended solids concentration of the feed sludge, the greater will be the production rate of the rotary vacuum filter (Figure 7-24), and the suspended solids concentration of the cake (Figure 7-25).

Generally, municipal wastewater treatment plant sludges are not concentrated beyond about 5-6 percent total solids (TS) [primary sludge + waste activated sludge (PS+WAS)], since above this concentration the sludge becomes difficult to pump, mix with the chemicals, and distribute after conditioning to the filter. Increasing production rates without higher sludge feed concentrations requires higher chemical dosages and results in higher cake moisture. Both of these consequences affect the cost of sludge dewatering and ultimate disposal.

The lowest feed sludge suspended solids concentration for successful vacuum filtration is generally considered to be 3.0 percent (PS+WAS). Below this concentration it becomes difficult to

Figure 7-24. Rotary vacuum filter productivity as a function of feed sludge suspended solids concentration.

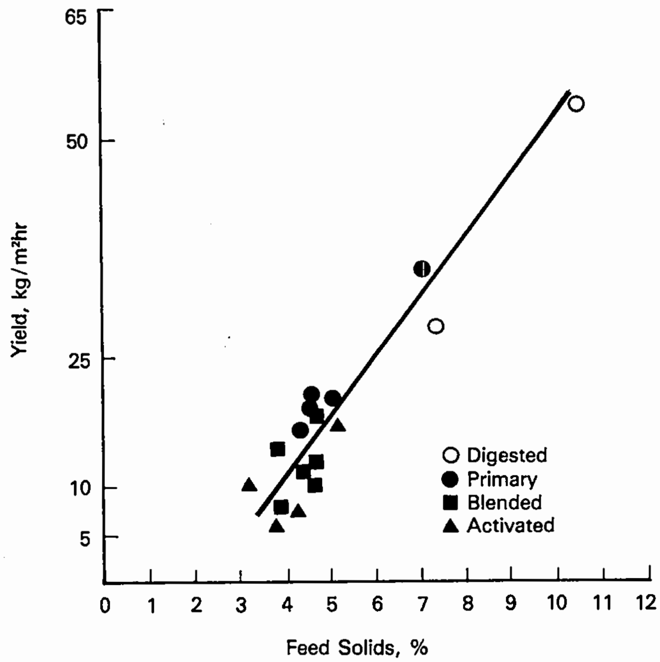
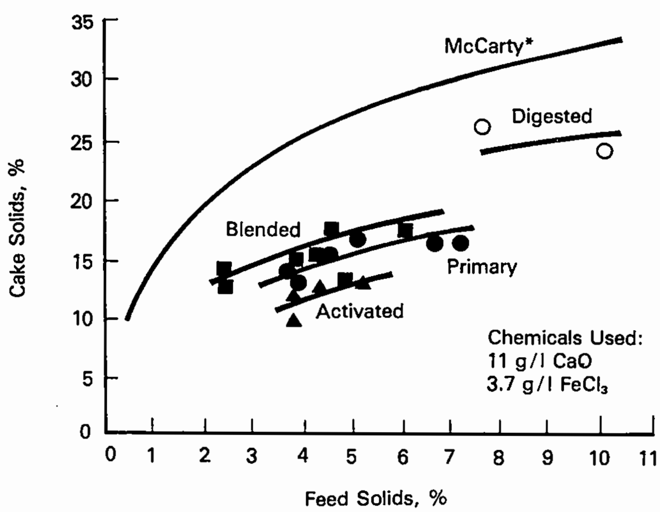


Figure 7-25. Sludge cake total solids concentration as a function of the feed sludge suspended solids concentration.



produce sludge filter cakes thick enough or dry enough for adequate discharge. For this reason, it is extremely important that the design and operation of the preceding sludge processes take into consideration the need for an optimal solids concentration when dewatering on vacuum filters. Existing operation can often be improved by increasing feed solids.

7.5.5.3 Performance Data

As with all types of mechanical dewatering equipment, optimum performance depends upon the type of sludge and its solids concentration, type and quality of conditioning, and how the filter is operated. Selection of vacuum level, degree of drum submergence, type of media, and cycle time are all critical to optimum performance. Tables 7-17 and 7-18 contain expected performance data for cloth and coil media rotary vacuum filters, respectively, for the sludge types indicated. Higher solids concentration (+3 to 4 percentage points) are produced by operating at about 50 percent of maximum rate.

Tables 7-19 and 7-20 contain operating data of several wastewater treatment plants using cloth media and coil media, respectively.

The efficiency of solids removal, or percent solids recovery, is the actual percentage of feed solids recovered in the filter cake. Solids removals on vacuum filters with adequate chemical conditioning range from about 85 percent for coarse mesh media to 98 percent with close weave, long nap media. The recycled filtrate solids impose a load on the treatment plant and should normally be kept to a practical minimum. However, it may be necessary to reduce the percent recovery in order to deliver more filter output and thus keep up with sludge production.

7.5.6 Equipment Selection Criteria

Due to high power costs and the heavy use of inorganic conditioners, vacuum filters are not often selected for use in new facilities. In some cases, refurbishing old equipment may be indicated to minimize capital costs. However, older units should generally not be used for standby capacity unless they are refurbished. The factors discussed in this section are included to provide guidance primarily for such applications.

Table 7-21 lists some of the advantages and disadvantages of vacuum filtration relative to other dewatering processes, and Table 7-22 lists design shortcomings that have been noted at a number of vacuum filter installations.

The significant points to be examined, if refurbishing or reusing vacuum filters, are:

- Media selection
- Feed solids
- Conditioning requirements and the design of the conditioning subsystem
- Sludge holding time before and after conditioning
- Filtration rate.

Other environmental considerations are shown in Chapter 4.

Table 7-17. Typical Dewatering Performance Data for Rotary Vacuum Filters - Cloth Media

Sludge Type	Feed Solids Conc.	Chemical Dosage[1], kg/Mg dry solids		Yield[2]	Cake Solids
	percent	$FeCl_3$	CaO	kg dry solids/m^2/hr	percent
Raw P	4.5-9.0	20-40	80-100	17-40	27-35
WAS	2.5-4.5	60-100	120-360	5-15	13-20
P+WAS	3-7	25-40	90-120		18-25
P+TF	4-8	20-40	90-120	12-30	23-30
Anaerobically Digested:					
P	4-8	30-50	100-130	15-35	25-32
P+TF	3-7	40-60	150-200	15-35	18-25
P+WAS	5-10	40-60	125-175	17-40	20-27
Elutriated Anaerobically Digested:					
P	5-10	25-40	0-50	20-40	27-35
P+WAS	4.5-8	30-60	0-75	15-35	18-25
Thermally Conditioned:					
P+WAS	6-15	0[3]	0	20-40	35-45

[1] All values shown are for pure $FeCl_3$ and CaO. Dosage must be adjusted for anything else.
[2] Filter yield depends to some extent on feed solids concentration. Increasing the solids concentration normally gives a higher yield.
[3] Some heat treated sludge requires some conditioning to maintain recovery at a high level.
1 lb/ton = 0.5 kg/Mg
1 lb/ft^2/hr = 4.9 kg/m^2/hr

Table 7-18. Typical Dewatering Performance Data for Rotary Vacuum Filters - Coil Media

Sludge Type	Feed Solids Conc.	Chemical Dosage[1], kg/Mg dry solids		Yield[2]	Cake Solids
	percent	$FeCl_3$	CaO	kg dry solids/m^2/hr	percent
Raw P	8-10	20-40	80-120	30-40	28-32
TF	4-6	20-30	50-70	30-40	20-32
P+WAS	3-5	10-30	90-110	12-20	23-27
Anaerobically Digested:					
P+TF	5-8	25-40	120-160	20-30	27-33
P+WAS	4-6	25-40	100-150	17-22	20-25
Elutriated Anaerobically Digested:					
P	8-10	10-25	15-60	20-40	28-32

[1] All values shown are for pure $FeCl_3$ and CaO. Dosage must be adjusted for anything else.
[2] Filter yield depends to some extent on feed solids concentration. Increasing the solids concentration normally gives a higher yield.
1 lb/ton = 0.5 kg/Mg
1 lb/ft^2/hr = 4.9 kg/m^2/hr

Equipment sizing may be accomplished by using information from Tables 7-18 through 7-20. This data may be augmented through the use of the filter leaf test as discussed in Chapter 5. A series of leaf tests will provide a range of values for solids loading and cake solids. The design conditions may be selected from these findings. A scale-up value of 0.8 is usually applied to obtain the final sizing. These procedures have been developed over many years and have excellent reproducibility and a high degree of confirmation.

7.6 New Methods of Dewatering

An extensive review of developments in municipal wastewater treatment plant sludge technology was undertaken for this section. This review included a literature search and contact with both commercial and non-commercial resources. Significant findings are identified and discussed below, including the Expressor Press, Som-A-System, CentriPress, Screw Press, and Sun Sludge System.

7.6.1 Expressor Press

A major manufacturer of dewatering equipment recently developed a modified twin belt press for use primarily in the industrial market. Substantial tests have been conducted with municipal sludges and various kinds of fibrous industrial waste sludges. The device, named the Expressor (R) or Expressor Press, consists of, in its basic form, two or three S rolls (wraparound) and a series of five P rolls (direct) on which the pressure can be individually varied. An Expressor Press with this configuration is shown in Figure 7-26.

Table 7-19. Specific Operating Results of Rotary Vacuum Filters - Cloth Media

Location	Sludge Type	Feed Solids Conc.	Conditioner Used[1]	Cake Solids	Yield	Filtrate
		percent	% by weight	percent	kg dry solids/m^2/hr	mg/l
Willoughby, Eastlake, OH	P + WAS + septic	4-6	$FeCl_3$ - 3 CaO - 14	20	14-24	
Tamaqua, PA	Anaer. Dig. (P + WAS)	6	$FeCl_3$ - 3 CaO - 23	18	15	SS 20-30
Grand Rapids, MI	Therm. Cond. (P + WAS)	10-15	None	50	30	SS 5,000 BOD 10,000
		3-4	$FeCl_3$ - 6 CaO - 16	19	15-17	
Grand Atkinson, WI	WAS	3.7	$FeCl_3$ - 8 CaO - 14	15	15	
Frankemuth, MI	WAS					
Oconomowoc, WI	Anaer. Dig. (P + WAS)	2.3	$FeCl_3$ - 6 CaO - 20	18	12-15	SS 500-1,100 BOD 10
Genessee City, MI	P + WAS	8	$FeCl_3$ - CaO- 16	27	27	

[1] All values shown are for pure $FeCl_3$ and CaO. Dosage must be adjusted for anything else.

Table 7-20. Specific Operating Results of Rotary Vacuum Filters - Coil Media

Location	Sludge Type	Conditioner Used[1]	Cake Solids	Yield
		% by weight	percent	kg dry solids/m^2/hr
Blytheville, AR	TF	$FeCl_3$ - 18 CaO - 47	33.1	50
York, PA	Anaer. Dig. (P + WAS)	$FeCl_3$ - 40 CaO - 125	21.1	23
Wyomissing Valley, PA	Anaer. Dig. TF	$FeCl_3$ - 31 CaO - 136	18.2	30
Bayonne, NJ	Anaer. Dig. P	$FeCl_3$ - 14 CaO - 120	30.9	38
Woodbridge, NJ	P	$FeCl_3$ - 20 CaO - 160	29.7	40
Shadyside, OH	Anaer. Dig.	$FeCl_3$ - 32 CaO - 165	29	20
Arlington, TX	TF	$FeCl_3$ - 32 CaO - 174	25.2	43

[1] All values shown are for pure $FeCl_3$ and CaO. Dosage must be adjusted for anything else.

In a second configuration, a unit called the Hybrid Expressor Press contains a gravity drainage section, four or five S rolls, and the five variable pressure P rolls. Depending on the model being considered, the P roll pressure can be varied from zero above the belt tension up to 200 kg/cm (1,000 lb/lineal inch). This new unit is capable of producing a very dry cake from the most difficult sludges with the use of press aids. A variety of press aids have been employed, but the most widely investigated material has been sawdust. The unit can produce an autogenous cake from waste activated sludge using between 50 and 125 percent sawdust by dry weight, based on the content of sludge solids. The water displacement by the press aid varies from slightly over one to as much as three kg H_2O/kg press aid added. The water displacement is based on the kg H_2O/kg sludge solids with and without press aid. The cake produced varies from 30 to 40 percent solids and, in some instances, runs somewhat higher than 40 percent.

Other press aids have been tested, including sand, soil, finely divided paper, fly ash, and coal fines. All work to some degree to increase the cake resistance to shear in the P rolls and hence permit higher

Table 7-21. **Advantages and Disadvantages of Vacuum Filtration**

Advantages	Disadvantages
Operation is easy to understand because formation and discharge of sludge cake are easily visible.	Consumes a large amount of energy per unit of sludge dewatered.
Does not require highly-skilled operator.	Vacuum pumps are noisy.
Will continue to operate even if the chemical conditioning dosage is not optimized, although this may cause discharge problems.	Lime and Ferric chloride conditioning can cause considerable maintenance cleaning problems.
Coil spring medium has very long life compared to any cloth medium.	The use of lime for conditioning can produce strong ammonia odors with digested sludge.
Has low maintenance requirements for a continuously operating piece of equipment, except in certain cases with lime conditioning.	Best performance is usually achieved at feed solids of 3-4%. However, some well conditioned sludges are filtered successfully at concentrations of <2%.
	Ferric chloride and lime conditioning costs are higher than polymer conditioning costs. Polymer conditioning is not always effective on vacuum filters.

Table 7-22. **Common Design Shortcomings of Vacuum Filter Installations**

Shortcomings	Resultant Problem	Solution
Improper Filter Media	Filter blinds, provides inadequate solids capture and/or poor cake release.	Replace media after testing for optimum.
Improper chemical conditioning used	Poor solids capture, low solids loading rate, and low cake solids concentration.	Change to correct chemical conditioners.
Inadequate water pressure for spray nozzles	Improperly cleaned media.	Provide booster pumping to maintain 484 kPa (70 psi) minimum pressure.

pressure and, in turn, higher solids content. Press aids in the 30 to 80 mesh region seem to be the most effective. With materials not particularly resistant to shear, such as paper and fiber, the particle size seems to have little impact on final sludge solids.

The press has also been tested on primary sludges and on mixtures of primary and waste-activated sludge from a pulp and paper manufacturing facility. Typical cake concentrations varied from 40 to 47 percent solids without a press aid. Wastes from the manufacture of pulp and/or paper would seem to work particularly well with this equipment because of the fibrous nature of the primary sludge.

Figure 7-26. **Expressor press.**

Also of interest is the ability of the press to produce an alum sludge cake of 40 to 60 percent solids using soil as a press aid in one test and sawdust in another. In each case, the press aid used was approximately 100 percent of the weight of dry solids of the alum sludge.

Determination of the pressure profile is a function of the sludge, the sludge blend, and the quantity and nature of the press aid used. On primary and waste activated sludges in the normal proportions (i.e., approximately 50-50) and on pure waste activated sludge, the P roll pressures are usually tapered and will vary from 10 kg/cm (56 lb/in) on the first roll to 60-250 kg/cm (336-1,401 lb/in) on the last roll.

The dewatered sludge shown in Figure 7-27 is from the Portland Columbia River Wastewater Treatment Plant, and was dewatered during a demonstration study at that facility. The activated sludge feed varied from 2.5 to 3.5 percent solids, and each test was run at approximately 100 percent of the press aid by weight. Sawdust additive yielded a cake in the range of 30 to 40 percent solids, while the paper press aid produced a cake from 35 percent to somewhat over 40 percent solids.

Solids capacity of the press varies from 225 to 600 kg/m hr (102 to 272 lb/m hr) and an acceptable hydraulic feed rate ranges from 1.6 to 3.2 l/s (25-51 gpm) on a 1-m (39-in) wide machine. The basic press has been investigated for further dewatering of cake derived from other dewatering equipment.

The press is of interest because it produces an autogenous cake with a modest amount of press aid. However, as of March 1987, there are no commercial installations on wastewater sludges. Initial units have been used in industrial and food processing applications. This device is similar to the high

Figure 7-27. Dewatered sludge at the Columbia River WWTP, Portland, OR.

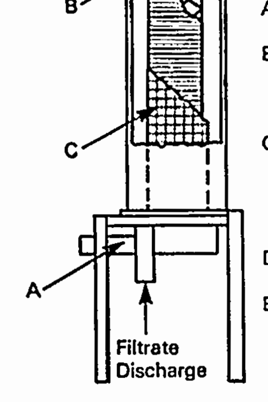
pressure attachment available on the Parkson Magnum 3000 and 3500 belt presses.

7.6.2 *Som-A-System*

The Som-A-System Screw Press consists of a vertical, rotating screw enclosed by dual stainless steel screens. The screens and screw are encased in a stainless steel housing with a removable cover on each side. Tiny perforations in the inner screen allow only water to escape. The outer screen has larger holes and easily collects the pressate, which sprays inside the housing and drains into a receptacle. Brushes are located along the edge of the screw to sweep the cake that builds up on the screen, allowing a clear opening for the pressate to escape.

The feed enters at the bottom of the screw press. A buildup of sludge cake on the screw is recommended to get good sludge dewatering. As the pressate drains, the cake becomes progressively drier and is pushed to the top, where it is discharged into a waiting dump truck or hopper. A back pressure system is located below the discharge chute and gives the cake a final squeeze before discharge. One plant, however, removed this cone, which collected hairballs, with no adverse affect to its operation or dewatering results (35). The Som-A-Press and Som-A-System are shown in Figures 7-28 and 7-29, respectively.

Figure 7-28. Functional schematic of Som-A-Press



A. Feed enters at inlet.

B. Variable speed auger carries liquid/solids stream vertically along the dewatering screen barrel.

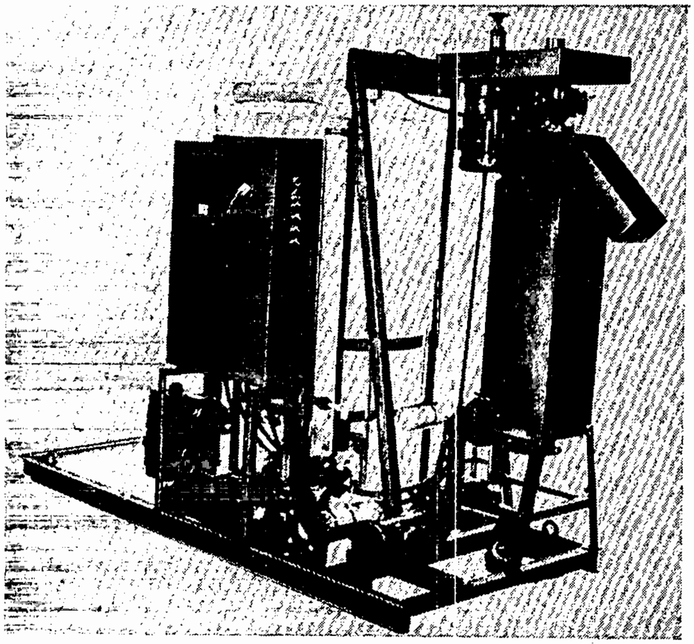
C. Most of the free liquid in the feed discharges through the screen openings in the initial 1/3 to 1/2 of the barrel length.

D. Solids build up, forming a plug.

E. Adjustable cone mounted on the auger shaft restricts passage of the cake and gives a final squeeze to the mass.

F. Revolving plug cutter pushes solids from the barrel into the discharge chute.

Figure 7-29. Som-A-System (courtesy of Somat Corp.)



Sludges that floc easily and are fibrous are the most conducive to a screw press operation. Feed concentration is critical to achieving high cake solids. The higher the feed concentration, the higher the cake solids and the unit capacity (kg/hr). Table 7-23 reports feed solids, cake solids, and solids recovery from several different plants using the Som-A-System (35). Key to the action of the unit is bridging of the holes in the screen, because the bulk of the particles in the sludge will be finer than the holes in the screen (36). Consequently, proper sludge conditioning is essential. Table 7-24 presents the polymer usage of several plants using the Som-A-System (35).

A slow screw speed will yield a better cake, although it will also decrease throughput. High flow rates and screw speeds generally result in a discharge of wet sludge. A variable speed pump regulates the feed rate to the screw press. At Pinetop, AZ the plant generally keeps the feed rate at 2.5 l/s (40 gpm), near the maximum. Sludge that has been aerobically digested at a 20 day detention will easily yield an acceptable 12-15 percent cake at a feed rate of 2.5 l/s (40 gpm). However, if the sludge had a lower detention time, a feed rate of 2.5 l/s (40 gpm) would produce a wetter cake.

Operation and maintenance of the Som-A-System is simple. Depending on the sludge, the press can normally be operated with only periodic checks. Many plants simply turn the machine on in the morning and periodically check the feed solids, the cake solids, and the level of the waiting dump truck. Some operations require more attention to the feed sludge

Table 7-23. Som-A-System Operating Data

Plant	Sludge	Average Plant Flow	Feed Solids TSS	Feed Rate	Cake Solids	Solids Recovery
		mgd	percent	gpm	percent	percent
Camden, NY	Aerob. Digested	0.6	1-2	10-24	10	85
Churchville, NY	Aerob. Digested	0.11	2.5-3.5	10.5	12.3	-
New Canaan, CT	Aerob. Digested	0.25	1.0-1.5	30-40	12-17	84-94
Danville, VA	WAS/Stab. scum	16.2	5-6	15	21-23	86
	from DAF		8	30-40	28-30	90
Pinetop, AZ	Aerob. Digested	0.4	2	40	12-15	88-90
Sunriver, OR	Aerob. Digested	0.5	0.5-0.75	35-40	7-12	85[1]
Frisco, CO	Aerob. Digested	1.0	2	15-18[2] 40[3]	11	-
Provo, UT	Anaer. Digested	1.5	2	30	7-15	87-94

[1] Normally the solids recovery runs 90-94%.
[2] Undersized polymer pump limits feed rate to 15-18 gpm - new pump ordered.
[3] With larger pump, expect to run presses at 40 gpm.

Table 7-24. Som-A-System Chemical Conditioning Data

Plant	Polymer			
	Name	Dosage	$/lb	$/ton DS
Camden, NY	Percol 767	5-20	-	-
Churchville, NY	Percol 757	1.57	2.80	4.39
New Canaan, CT	Percol 757	6-9	3.25	24.38
Danville, VA	Cationic	21	0.92	19.32
Pinetop, AZ	Percol 757	8-10	-	27.00
Sunriver, OR	Allied CC4450	21	1.65	34.65
Frisco, CO	Percol 757	12	2.70	32.40
Provo, UT	Percol 763	21	-	-

to ensure that the proper concentration - and not water - is being fed to the press. The unit is also relatively easy to disassemble. General maintenance involves routine lubrication and washing the screens to prevent buildup of sludge, which can prematurely wear the brushes. Repairs reported by plants have been limited to replacement of inner and outer screens and brushes. One plant replaced the brushes after approximately 1,500 hours (35).

The low capital cost of this screw press is a primary attraction and comparative economic evaluations point favorably to it. For one plant, the Som-A-System was approximately $55,000 less than a belt filter press bid for the same job. It is ideal for operations with limited space requirements, since the system occupies, at a maximum, approximately 3 m^2 (32 sq ft) of floor space.

Potential drawbacks include low unit capacity, higher polymer dosage, and lower cake solids. Capacity of the presses can be a deterrent because the small throughput demands a multiplicity of units, which can be more difficult to control. A few plants (35) expressed disappointment about the amount of polymer required, and were experimenting in an effort to reduce the quantity.

7.6.3 CentriPress

Based on observed field demonstrations at the 1987 IFAT Conference held in Munich, Germany, there have been significant improvements in the capabilities of a newly designed solid bowl continuous flow centrifuge. The improvements were in the area of cake solids concentration. In testing, the centrifuge was operated in parallel with a filter press system. The new centrifuge design, called the CentriPress, produced as a high a cake solids as the filter press system. Figure 7-30 shows the CentriPress and examples of cake that it produced.

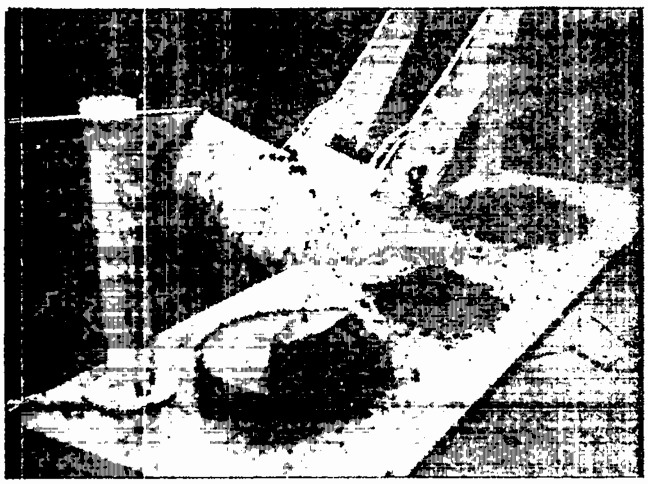
A Model S2-1, 45-cm dia. x 135-cm long (18-in x 53-in), centrifuge at the Marienfelde STP was operating on digested primary and waste activated sludge. This same sludge was fed to 91.5-cm diameter x 274-cm long (36-in x 108-in) centrifuges which were dewatering the plant sludge to a cake product of approximately 22 percent TS. The CentriPress was producing a granular cake of 30-32 percent TS. The "standard" centrifuges produced a cake with a 60 percent higher moisture content. Both centrifuge installations were recovering in excess of 90 percent of the feed solids and the products are shown in Figure 7-30.

A larger unit, 91.5 cm dia. x 274 cm long (36 in x 108 in), is operating at Vienna, Austria WWTP. This unit is dewatering a heated primary and waste activated sludge to a cake solids content of 40-42 percent TS. Results from the centrifuge are comparable to those produced by a recessed plate filter press.

The manufacturer has advised that orders have been taken in Europe for the new machine, and that demonstration of the centrifuge capabilities were initiated in the spring of 1987 in the United States. Details of the centrifuge construction were not made available prior to printing.

Figure 7-30. CentriPress and cake samples.

Humboldt-Wedag Centri Press



From Left to Right, Pressate, Press Cake, Ash Recycle, Centrifuge Cake, and Centrate

Granular Centrifuge D (P + WAS) Cake @ 30-32% TS

Test trials using a Humboldt-Wedag CP2-1 were performed by the Metropolitan Sanitary District of Greater Chicago (MSDGC) at the West-Southwest STP. This plant currently employs high-speed centrifuges for dewatering a digested primary and waste activated sludge, which has an original solids ratio of 0.21 PS: 0.79 WAS. The existing centrifuges produce a cake of 14-16 percent TS.

The tests were conducted using two types of cationic polymers as shown in Table 7-25. One of the polymers was not cost effective for the plant's digested sludge. The tests used different feed rates and differential speeds, with the polymer adjusted to maintain the TSS recovery in the range of 85-95 percent. The key results of Table 7-25, using American Cyanamid 2540C polymer are as follows:

	Average	Range
Cake Solids, %	29.4	26.2-33.9
Solids Recovery, %	92.7	78.4-97.9
Polymer Dosage, kg/Mg	7.45	3.23-15.93
$/Mg	3.37	5.86-29.22

Figure 7-31 shows the effects of polymer dosage on the solids recovery of the CentriPress. About 5 kg/Mg (10 lb/ton) of cationic polymer was required to maintain the solids recovery in excess of 90 percent TS. Table 7-25 does not indicate that higher dosages of polymer were beneficial to improve cake solids, although recoveries above 95 percent were achieved.

The use of low differential speeds appears to be the key to achieving good cake solids. As shown in Figure 7-32, there was a good correlation between cake solids and centrifugal force at about 2,600 g's.

7.6.4 HIW Screw Press

This Korean screw press is being evaluated for dewatering sludge from liquid to cake, Second stage (cake to drier cake) operations have also been evaluated. The HIW screw press, shown in Figure 7-33, is continuously fed a polymer conditioned sludge. Once inside the unit, the sludge receives a gradually increasing pressure as it progresses through the screw press. The maximum pressure before discharge may exceed 10 kg/cm^2 (147 lb/sq in). In some instances the dewatering may be enhanced by heating (a normal experience with screw presses) prior to the dewatering screw. This screw press is said to be relatively simple and easy to maintain. Also, the low operating speed helps keep repair costs to a minimum. HIW reports that there are over 100 units in operation (or installed) for various types of wastewater treatment and are providing satisfactory service. An adequate U.S. database is not yet available. Some results reported are shown in Table 7-26.

More data and a better definition of the feed sludge is required to fully evaluate the possibility of the screw press replacing conventional dewatering equipment. Past excessive secondary solids losses must be evaluated as a function of the cake solids content produced.

During May of 1986, the Municipal Sanitary District of Greater Chicago (MSDGC) tested a pilot HIW screw press. The unit was tested on primary and anaerobically digested sludge at the West-Southwest STP. The test was performed over a period of two days and approximately twelve separate runs were undertaken. Sludge flow rate, dilute polymer concentration, and polymer flow were varied. With an average sludge feed concentration of 4.5 percent, the test unit attained the following average results.

Cake concentration:	17.5%
Solids recovery :	94.5%
Pressate concentration:	3,720 mg/l (0.43 lb/gal)
Polymer usage:	8 dry kg polymer/dry Mg solids (16 lb/dry ton)

Based on these pilot-test results, the MSDGC decided to purchase a full-size screw press, to be delivered in mid-1987. MSDGC anticipates that a full-size screw press may be a cost-effective alternative to centrifugation due to the following considerations:

- Low initial cost
- Lower electric power consumption
- Equal to or higher cake concentrations
- Slow operating speed (low G force)
- Lower maintenance cost
- Comparable polymer cost.

7.6.5 Sun Sludge System

The Sun System (Hi-Compact) of pressing sludge was developed in Japan, and has been licensed for marketing and manufacture for Europe and the United States. The principle of the process is to develop a structured material from a cake of poor dewatering characteristics, and to form liquid channels. The cake is then subjected to high pressures. To that end, dewatered sewage sludge is reduced to pellets which are subsequently coated by a powdery layer of a drainage substance such as ash, pulverized coal, etc. Compressing a stack of these pellets results in a compact block interwoven with a network of drainage layers; the water being removed by pressing flows through a line of least resistance to the nearest drainage layer as shown in Figure 7-34.

In the system, sludge is first dewatered by conventional dewatering equipment such as vacuum filters, centrifuges, or continuous belt filters to a 20-25 percent solids concentration. This material is then conditioned in a unit called a disintegrating pelletizer,

Table 7-25. Results of Chicago WSW CentriPress Study

Run	Machine Data		Sludge Data				Polymer Data		Performance			
	G-Force, g's	Diff. Speed	Feed Rate, gpm	Feed Conc., %	Feed Solids, ton/d	% Volatiles	Flow Rate, gpm	Polymer (Dry), lb/ton	Cake Solids, %	Centrate Solids, %TSS	Capture, %	Cost, $/ton
1[1]	2,300	3	27	4.18	6.78	48.3	3.45	14.67	29.1	4,200	91.27	11.24
2	2,300	2	27	4.19	6.79	48.0	3.09	13.12	26.3	1,700	96.57	10.06
3	2,300	1.8	25	4.00	6.01	47.0	3.28	15.73	29.6	5,000	89.00	12.06
4	2,300	2	31	4.09	7.61	46.7	4.10	8.41	29.2	1,000	97.89	6.45
5	2,600	2	26.5	4.16	6.62	48.4	3.9	10.61	33.2	3,200	93.21	8.74
6	2,600	2	26.5	4.10	6.52	48.9	3.9	10.28	33.9	2,700	94.17	8.88
7	2,600	3.5	32	3.96	7.61	48.7	3.73	9.42	29.7	7,200	83.85	7.76
8	2,600	5.5	32	4.07	7.82	49.0	2.63	6.46	26.2	10,000	78.42	5.33
9	2,600	5.5	32	4.10	7.89	49.2	4.58	11.15	27.4	4,200	91.15	9.19
10	2,600	2.2	16	4.12	3.96	48.6	3.27	19.83	30.2	1,900	95.99	16.34
11	2,600	2.8	16	3.80	3.65	50.1	2.67	17.57	31.1	2,400	94.41	14.48
12	2,600	7.5	50	4.13	12.43	48.8	5.57	10.76	29.8	1,400	97.07	8.87
13	2,600	5.8	50	4.11	12.34	48.1	5.57	11.93	28.4	3,700	92.20	9.83
14	2,600	2.5	22	4.23	5.59	50.7	3.91	21.84	28.5	2,900	94.10	18.20
15	2,600	2.5	22	4.25	5.62	50.5	4.4	24.45	28.4	1,500	96.98	20.37
16	2,600	2.7	22	4.20	5.55	50.1	3.2	31.85	28.9	1,300	97.34	26.56
17[2]	2,600	2.5	22	4.25	5.62	49.4	2.73	33.25	29.2	2,900	94.11	76.48
18	2,600	2.8	22	4.25	5.62	49.4	2.73	33.25	32.0	1,700	96.51	76.48
19	2,600	2.9	22	4.00	5.28	49.8	2.4	31.12	29.4	1,100	97.61	71.58
20	2,600	2.2	22	4.10	5.42	49.2	2.4	30.31	31.3	2,200	95.30	69.71
21	2,600	2	22	4.08	5.39	50.6	3.57	45.34	32.8	1,300	99.20	104.28
22	2,600	2	22	4.16	5.50	50.5	3.50	42.8	34.8	2,200	95.31	98.44
23	2,600	5	26.5	4.09	6.51	49.3	3.50	36.16	30.0	2,400	94.84	83.17
24	2,600	1.5	18	4.05	4.38	49.1	2.5	38.39	30.8	1,000	97.85	88.30
25	2,600	1.2	18	4.08	4.41	49.4	2.9	44.23	29.7	1,200	97.45	101.73

[1] Tests 1 through 16 used American Cyanamid 2540C polymer.
[2] Tests 17 through 25 used Allied Chemical Percol 778F525 polymer.

which first breaks and forms the sludge cake into small particles and then coats the particles with a dry powder, forming sludge-like pellets. The dry additive used should be mostly water insoluble and should not break up at the high pressures used. Materials such as diotomaceous earth, gypsum, calcium carbonate, incinerator ash, coal powder, bone meal, dried pulp, sawdust, and soil have been used, either alone or in combination with each other. The conditioners should be added in the ratio of 40-60 percent by weight per unit dry weight of the original sludge cake. The effective sludge particle or pellet's diameter should not be greater than 20 mm (0.8 in). Best performance occurs when the effective diameter of the pellets is between 3 and 5 mm (0.1-0.2 in). Also, the conditioning agents should coat only the surface and should not be kneaded into the sludge pellets for maximum effectiveness.

The conditioned sludge cake particles are conveyed to a hydraulic press where additional water is removed, and a cake of greater than 40 percent solids is produced. The pelletized sludge is pressed between two sheets of filter cloth that cover thick plates that have a number of perforations 2-10 mm (0.1-0.4 in) in diameter. Compression is carried out in two steps. The initial compression step is usually at 15-25 kg/cm^2 (210 lb/sq in) for 45 seconds followed by a pressure of 30 kg/cm^2 (430 lb/sq in) for 5 minutes. In practice, the compression has occurred at 15 kg/cm^2 (210 lb/sq in) for 45 seconds, followed by a pressure of 30 kg/cm^2 (430 lb/sq in) for one minute. The pelletized sludge cake is compressed by an oil hydraulic cylinder to form a disc-shaped solid with a 40-55 percent solids concentration. As an example, a mixture of primary and waste activated sludge having a 2 percent solids concentration could first be dewatered by a belt filter press to a solids concentration of 25 percent and then, with the Sun Sludge System, could be further dewatered to a solids concentration of 55 percent.

Figure 7-31. Effect of polymer dose on solids recoverv.

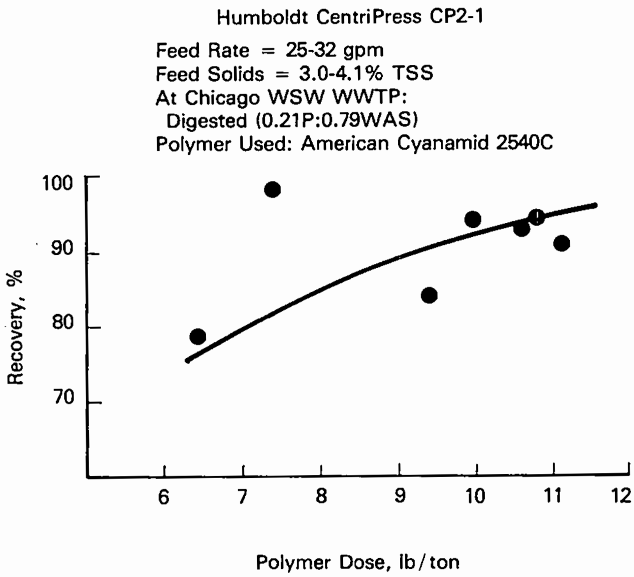
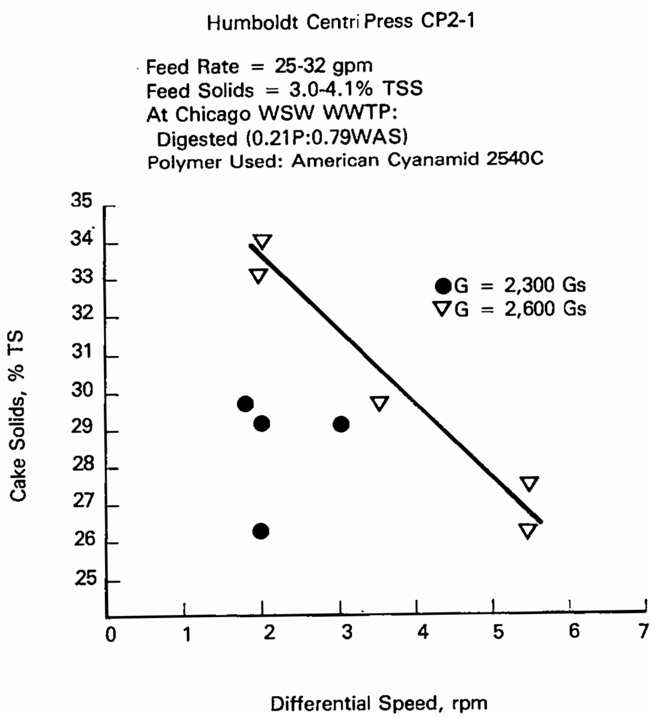


Figure 7-32. Cake solids vs. differential speed.



The Ashigara Works of Japan has successfully been using this process for waste activated sludge treatment since mid-August 1982. The excess sludge is dewatered by a belt press to a water content of 80 percent, then pelletized and conditioned with incinerator ash and further dewatered to a water content of 50 percent or less. Ash is added in the ratio of 10 to 15 percent by weight of the amount of belt press cake (or 50-75 percent of the dry solids). Sludge cake is incinerated and heat recovery equivalent to 30 l/hr (7.9 gal/min) of fuel additive is practiced.

At the 1987 IFAT show in Munich a field demonstration of the process produced a cake of 55 percent from a 32 percent sewage cake mixed with sludge ash (50 percent by dry solids weight) every 3 minutes. The unit was pilot scale producing in excess of 1,000 kg cake/hr (2,204 lb/hr). This would be equivalent to 370 kg/hr (816 lb/hr) of sewage sludge solids. In this demonstration, pressures up to 60 kg/cm^2 (853 lb/sq in) were employed and the press time was shortened to about three minutes.

While the product from the press is very hard, it is also quite friable. It can be easily fragmented into particles which are dry to the touch and can easily be transported pneumatically. The pelletizing/pressing operation at Munich reduced the moisture content from 3.5 kg H_2O/kg TS to 1.3 kg H_2O/kg TS (sludge only basis). The feed would have been suitable for boiler feed and would produce an equilibrium temperature of about 1,090°C (2,000°F).

While the process mechanics look favorable, the machine design capable of long-term operation at 50-60 kg/cm^2 (710-850 lb/sq in) will need to be further evaluated.

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Figure 7-33. Schematic of the screw press dewatering system (courtesy of Hoilin Iron Works [HIW]).

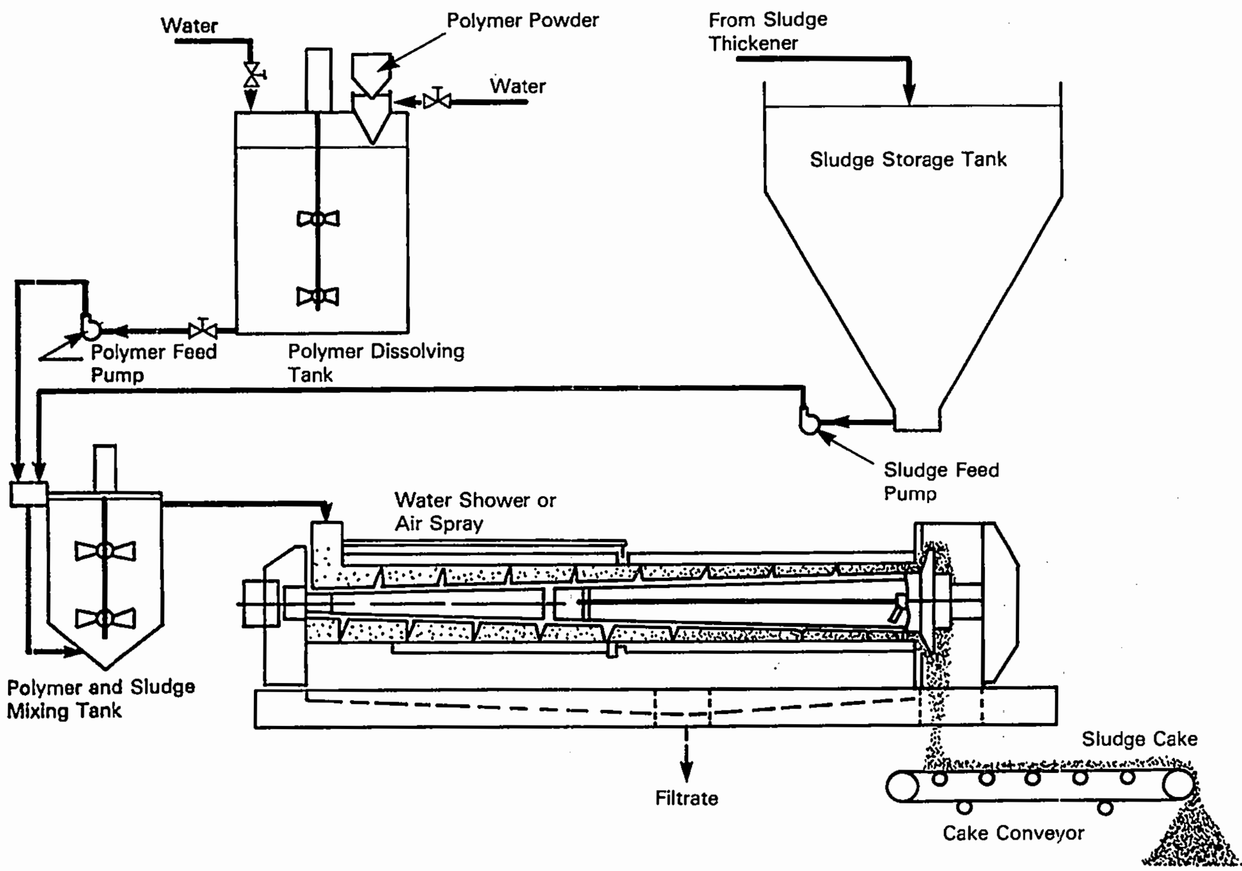

Table 7-26. Test Results for HIW Screw Press

Sludge	P/S Ratio	Feed Solids	Cake Solids	Solids Recovery	Polymer Dosage
		percent	percent	percent	kg/Mg
Digested A	10/90	4.65	20.9	93.0	9.1
Digested B	10/90	4.93	25.3	97.8	7.6
Primary A	NR	2.85	20.5	95.0	13.4
Primary B	NR	2.37	21.2	95.9	16.0
Paper Mill 1	0/100	3.45	48.6	99.0	1.0
Paper Mill 2	60/40	4.08	44.6	98.9	NR
Paper Mill 3	50/50	2.95	42.3	98.9	NR
Paper Mill 4	0/100	2.4	23.0	95.4	NR

[1] Not reported.

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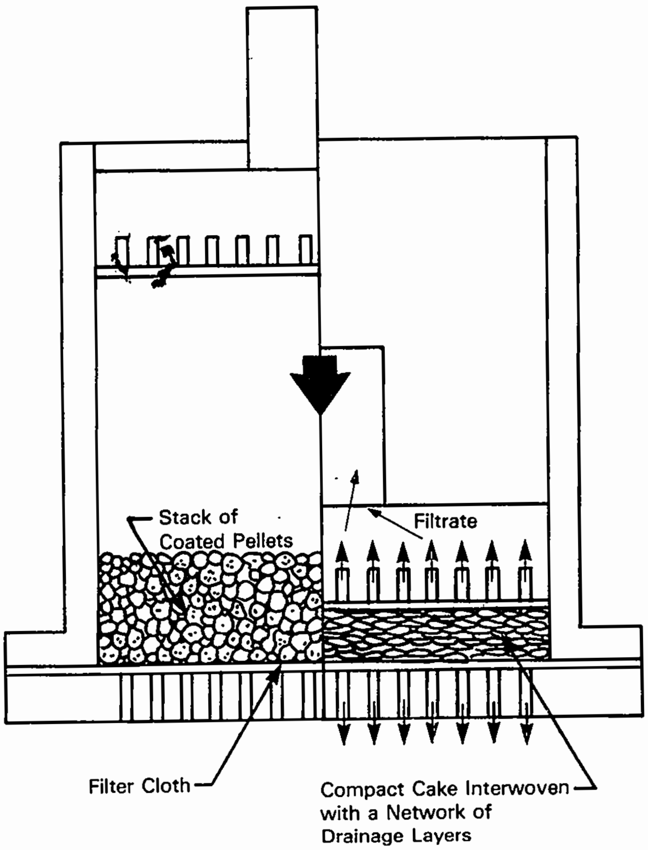
Figure 7-34. Sludge press function of the Hi-Compact method (courtesy of Humboldt-Wedag).



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Chapter 8

Case Studies: Air Drying Systems

8.1 Introduction

Two air drying technologies are described in this chapter: upgraded sand drying beds that use polymer to improve dewatering and paved drying beds that use a composter/agitator to increase the rate of dewatering.

8.2 Upgraded Sand Drying Beds

A site visit was conducted to write the case study for an upgraded sand drying bed at Elgin, Illinois. Information on three other upgraded sand drying beds was obtained through personal communication.

8.2.1 Detailed Case Study - Elgin, Illinois

The Sanitary District of Elgin, Illinois operates three wastewater treatment plants. The North Plant is a 0.15-m^3/s (3.5-mgd) activated sludge plant. Following is a description of the plant as it operated in 1986.

The activated sludge process is a conventional, complete-mix design with surface mechanical aerators. The plant has an average daily flow rate of 0.14 m^3/s (3.15 mgd) and is undergoing a construction expansion to 0.25 m^3/s (5.75 mgd).

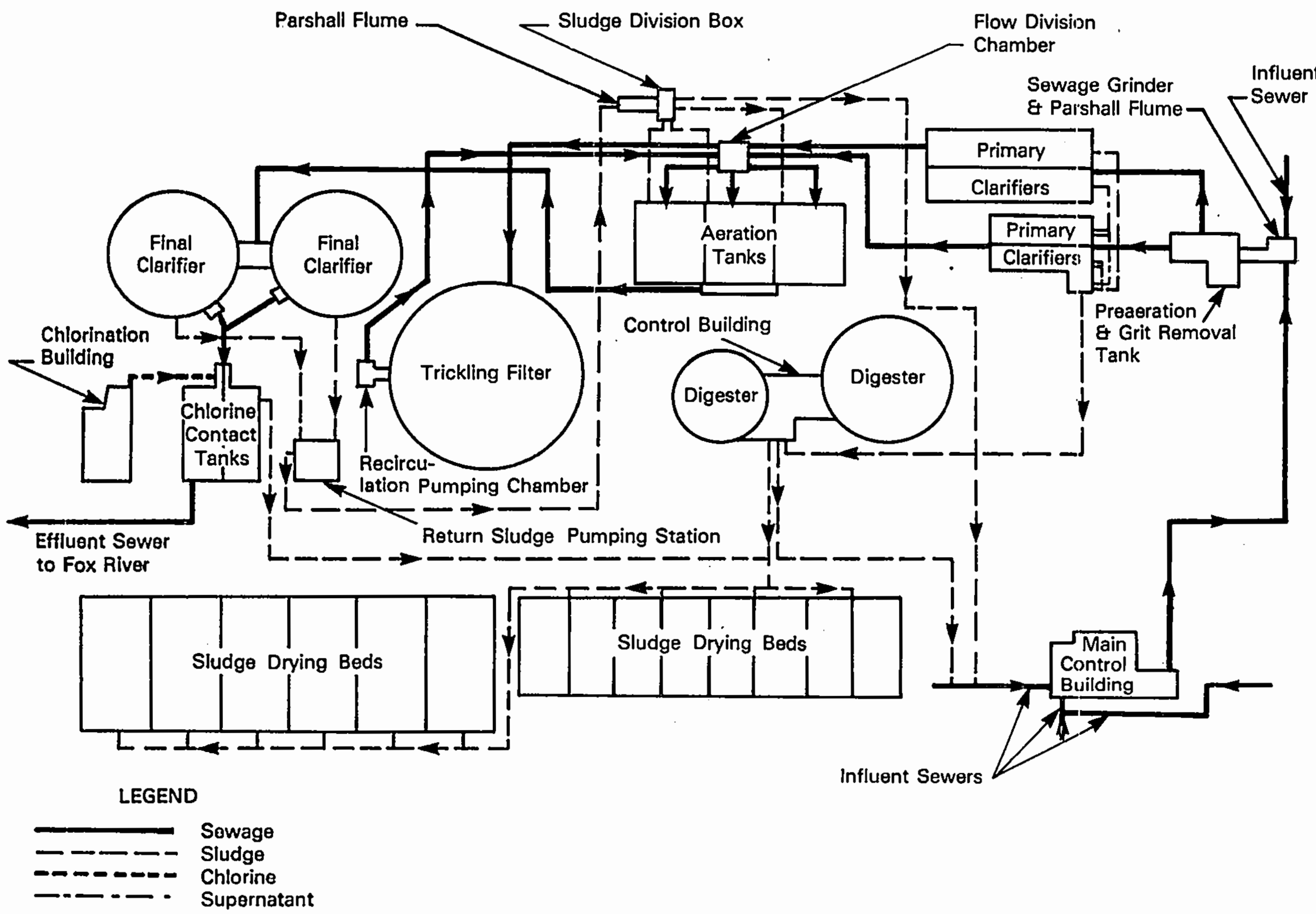
The waste activated sludge (WAS) at the North Plant is wasted to the primary clarifiers. The combined primary sludge (P) plus WAS is digested in two-stage anaerobic digesters. Following digestion, the sludge is dewatered on sand drying beds and the dewatered sludge is hauled away to various land application sites. A plant schematic for the North Plant is shown in Figure 8-1.

Polymer is added in-line to the sludge as it flows from the primary digester to the secondary digester. The polymer serves as a settling aid during supernatant removal operations. Using polymer enables the District to remove more supernatant and produce a thicker digested sludge. The polymer dosage applied to the digester is about 114 liters (30 gal) of liquid cationic polymer to 454 m^3 (120,000 gal) of sludge, four times per month. The approximate polymer dosage is 19.5 g/kg (39 lb/ton) of solids, or about $7.16 per Mg ($6.50/ton) of solids. The first stage digester is both heated and mixed, while the second stage digester is neither heated nor mixed. The digested P + WAS has an average solids concentration of 5 to 6 percent.

Polymer is also added to the sludge as it flows onto the drying beds. The approximate polymer dosage is 32 g/kg (65 lb/ton) of solids, or about $12.04/Mg ($10.92/ton) of solids. (This polymer dosage is higher than average due to inefficient mixing, as described below.) Polymer addition to the sand drying beds and to the anaerobic digesters has been practiced for at least 8 years. The Sanitary District of Elgin first constructed the North plant in 1962 to a design capacity of 0.035 m^3/s (0.8 mgd). In 1972 the plant was expanded to a design capacity of 0.11 m^3/s (2.4 mgd). Later the plant's average design capacity was upgraded on paper to 0.15 m^3/s (3.5 mgd). By the mid-1970s the plant, during the spring of each year, produced more sludge than could be dewatered. This was the time when polymers were first tested to aid dewatering on the sand drying beds. Soon thereafter, the use of polymer became a common practice. The main reason for its use is that the sludge dewaters easier and faster with polymer. The total dewatering time required is usually 3 to 4 weeks during the non-winter months. If polymer is not added to the sludge, the dewatering time is typically 5 to 6 weeks. Sludge is applied to the beds year-round even though the sludge freezes fairly often during winter months.

As the digested sludge flows toward the drying bed and discharges onto a concrete splash pad, liquid polymer is added to the sludge through a separate hose that discharges onto the same splash pad. (This is an unusual and inefficient method for polymer addition. It is more typical and more efficient for polymer to be added in-line, upstream of the drying bed.) The concrete splash pad and sludge feed line are located between two sludge beds. Gates on either side of the splash pad can be raised or lowered to allow sludge to flow onto either the right or left sludge bed. Liquid sludge is added to the drying bed to an initial depth of about 23 cm (9 in). During warm summer months, a dewatered sludge with a solids concentration of between 28 to 32 percent solids can be removed in 3 to 4 weeks. During spring and fall months, the dewatered sludge solids concentration

Figure 8-1. Schematic of the North Plant, Elgin, IL.



can be as low as 20 percent. During winter months, sludge has occasionally been removed from the beds in a frozen state.

Dewatered sludge is removed from the drying beds with a front-end loader. The loader can easily remove the sludge because the sand bed has concrete strips that run lengthwise down the entire length of the bed. The concrete strips are 0.7 m (2 ft, 3 in) wide, 20 cm (8 in) thick, and are separated by open sand strips 0.8 m (2 ft, 9 in) wide. Vitrified clay tile underdrains are located beneath the sand strips. Sand bed filtrate is returned by gravity to the plant headworks.

Following mechanical removal of most of the dewatered sludge, small quantities of sludge remaining in the sand, alongside the concrete strips, or at the edge of the beds are removed manually with a rake. This step prevents clogging of the sand filtering media. The sand is then loosened as needed with a rake to allow better percolation through the sand. If the sand is not loosened, it becomes packed down, reducing percolation. Following raking of the sand strips, the entire bed surface is covered with about a 2.5-cm (1-in) layer of sand. The sand is scraped off the concrete strips and is used to fill the void areas created during the last cleaning of the bed.

Because of the construction of the current plant expansion, the total number of beds available for sludge application was reduced from 14 to 7 starting in August of 1984. The total sludge bed area before the reduction of bed area in 1984 was about 2,750 m^2 (29,600 sq ft), or an average of 116 m^2 (2,114 sq ft) per bed. In the period of August 1983 through July 1984, the average daily wastewater flow rate was 0.13 m^3/s (3.1 mgd). The sludge production was 3,230 m^3 (854,000 gal) at an average digested sludge solids concentration of 5.6 percent. Thus, the digested sludge production is about 1,350 dry Mg of solids/yr/m^3/s (65.4 dry tons/yr/mgd) of average plant flowrate. This amounts to only 3,710 kg solids/d/m^3/s (358 lb/d/mgd). One reason for such a low amount is that the influent sewage strength is quite weak; during the period of August 1983 through July 1984, the average influent BOD_5 and suspended solids concentrations were 98 mg/L and 126 mg/L, respectively.

The District Engineer cited industrial waste contributions as one of the problems at the North Plant. For the activated sludge system, the average F/M ratio was 0.6 and the average aeration basin hydraulic detention time was 3.9 hours during 1983-84. The annual average solids loading rate applied to the North plant sludge beds during the year before the reduction to 7 beds, based upon an average wastewater flow of 0.13 m^3/s (3.1 mgd), is 66.4 kg solids/m^2/yr (13.6 lb/sq ft/yr).

The polymer system was installed by plant operators at a minimal construction cost. The operation and maintenance cost for the drying beds is broken down in Table 8-1, which describes the cost to fill and clean one average size bed.

During the period of August 1983 through July 1984, the total weight of dry sludge solids remaining following anaerobic digestion was 183 Mg (201 tons). At an average digested sludge solids concentration of 5.6 percent, the total volume of sludge dewatered was 3,230 m^3/yr (854,000 gal/yr). At a depth of 23 cm (9 in) of sludge and an average sludge bed area of 116 m^2 (2,114 sq ft), the average volume of sludge applied per bed is 45 m^3 (11,900 gal). This volume of sludge results in a total number of sludge applications to a drying bed of 72 per year, or 5.1 applications per year per bed. This total is low because of the low-strength influent wastewater and low sludge production; the sludge beds were not used at their full capacity in 1983-84. Based upon an average cost of $135.95 per sludge bed application (from Table 8-1), the total annual cost during 1983-84 was $135.95 x 72, or $9,800/yr.

Currently, with about two-thirds of the North plant sludge beds removed, only a portion of the sludge can be dewatered on the sand drying beds. Dry sludge is hauled away by a contract hauler to various land application sites for a 1986 price of $7.52/$m^3$ ($5.57/cu yd) of dry sludge. Assuming the dewatered sludge cake has a solids concentration of 30 percent and a specific gravity of about 1.15, the cost of hauling dry sludge is $21/Mg ($19/ton) of dry solids. Undewatered liquid sludge is hauled away to land application sites for 0.77 cents/l (2.9 cents/gal) of liquid sludge.

8.2.2 Case Study - Belleville, Illinois

The following case study describes sludge dewatering at the wastewater treatment plant for Belleville, Illinois as it operated in 1986. The plant, located about 32 km (20 mi) southeast of St. Louis, is a 0.4-m^3/s (8-mgd) activated sludge wastewater treatment plant. Current average daily flows are about 0.28 m^3/s (6.5 mgd). Waste activated sludge is returned to the primary clarifiers and the combined sludge is pumped to anaerobic digestion. Sludge treatment at the plant consists of two-stage anaerobic digestion and dewatering on sand drying beds or storage in lagoons.

There are 11 sand drying beds and a total sludge bed area of 11,700 m^2 (125,900 sq ft). A liquid cationic polymer is added to the digested sludge before it flows onto the drying beds. The typical polymer dosage is 11 g of liquid polymer/kg (22 lb/ton) of solids, or about $11.62/Mg ($10.54/ton) of solids. The polymer is pumped into the sludge discharge line from the anaerobic digesters at a point about 244 m (800 ft) before the sludge beds. Digested sludge is normally applied to the beds at a solids concentration of 3 to 3.5 percent and at a depth of 36 to 38 cm (14 to 15 in).

The sludge beds are used regularly from March through October. During the winter months, the sludge is only occasionally applied to the beds since dewatering has nearly ceased. The sludge takes normally about 6 to 8 weeks to dry with the polymer and about 12 weeks without polymer. Application of polymer to the sand drying beds has been practiced at Belleville since 1965. The Wastewater Treatment Division Superintendent considers it more cost-effective to dewater sludge on sand drying beds with polymer due to the reduced drying time provided by the polymer.

The dewatered sludge is removed from the drying beds at about 30 to 35 percent solids. Sludge is removed from the drying beds manually with shovels. Two men shovel sludge from the bed into a truck for transfer to a stockpile site. Stockpiled sludge is removed from the plant site for use by farmers, gardeners, nurseries, etc. Sludge removal from one bed takes about 24 man-hours. Cost estimates for the sludge bed operations for 1985, prepared by the Wastewater Treatment Division Superintendent, are presented in Table 8-2. The total operation and maintenance cost estimate is $21,236/yr, or $56.55/dry Mg ($51.30/dry ton) of solids. The solids loading rate during 1985 is 54 kg/m^2/yr (11 lb/sq ft/yr).

8.2.3 Gwinnett County Water Pollution Control Department, GA

The Gwinnett County Water Pollution Control Department, located about 40 km (25 mi) east of downtown Atlanta, operates 11 wastewater treatment plants. Polymer addition to improve dewatering on sand drying beds is practiced at three of the plants. The design capacities of the three plants are 0.20, 0.13, and 0.09 m^3/s (4.5, 3, and 2 mgd). The sludge bed area at these plants is 2,900 m^2 (31,000 sq ft) for 9 beds, 2,000 m^2 (21,000 sq ft) for 6 beds, and 930 m^2 (10,000 sq ft) for 3 beds, respectively.

Table 8-1. Elgin, IL North Wastewater Treatment Plant Operation and Maintenance Costs for One Drying Bed Cycle (average bed area = 196 m^2 (2,114 sq ft)

Item	Operation and Maintenance Costs Units	Unit Cost	Cost
Labor - Add sand Add sludge Add polymer	2 hr	$15/hr	$30.00
Labor - Clean bed Haul dewatered sludge	2 hr	$15/hr	$30.00
Electricity - 2.2-kW (3-hp) sludge feed pump 0.25-kW (0.33-hp) polymer pump	0.12 MJ (3.3 kWh)	$1.67/MJ ($0.06/kWh)	$0.020
Fuel, diesel - for loader, haul truck, and sand replacement	57 l (15 gal)	$0.22/l ($0.85/gal)	$12.75
Polymer - 79 l/45,000 l sludge, 1.01 kg/l (8.47 lb/gal)	81 kg (178 lb)	$0.37/kg) ($0.168/lb)	$30.00
Sand, replacement	8,845 kg (19,500 lb)	$3.73/Mg ($3.38/ton)	$33.00
Total Cost per Bed			$135.95

Table 8-2. Sludge Bed Operation - Belleville, IL, 1985

Eleven (11) Drying Beds - 125,900 total sq ft

Estimate Costs	
Bed filling: 28 beds x 8 hr = 224 x $14 =	$3,136.00
Stripping: 28 beds x 24 hr = 672 x $13 =	$8,736.00
Sand - Power-Vehicle, etc.	$5,000.00
Polymer: $10.54/dry ton x 414 tons =	$4,364.00
Total	$21,236.00

$21,236 ÷ 414 dry tons = $51.30/dry ton

The sludge beds were constructed in 1950, and the polymer system in use at this time was purchased in 1967. The vehicle used for stripping the beds is a 1971 Chevrolet pickup truck that was modified for sludge bed use at a cost of $2,500 in 1984.

All three wastewater treatment plants are activated sludge plants with single-stage nitrification and no primary clarifiers. The two largest plants also remove phosphorus with the addition of alum to the secondary clarifier. All sludge is aerobically digested. Sludge is typically removed from the aerobic digester at about 4 percent solids for the two largest plants and at 2 percent solids at the smallest plant. At two of the three plants, a liquid cationic polymer is pumped into the digested sludge line as sludge flows to the drying bed. At the third plant, a crude mixing box is used to allow the sludge to mix with the polymer. The polymer dosage is 5 to 6.5 g liquid polymer per kg (10 to 13 lb/ton) of solids, at a cost of $8-10 per Mg ($7-9/ton) of solids.

Sludge is applied to the sand drying beds year-round. The approximate depth of sludge applied is 10 to 13 cm (4 to 5 in) per application. In the summer the typical drying time is 2 to 3 weeks. The sludge surface on the beds normally cracks overnight when using the polymer. Without polymer, the typical drying time is 4 to 5 weeks, and the sludge surface takes 2 to 3 days before cracking. The solids loading rate achieved at each of the three plants using polymer on the sand drying beds ranges from 200 to 220 kg/m^2/yr (41 to 44 lb/sq ft/yr).

Polymer has been applied to the sand drying beds for 2 1/2 years at the largest plant, 1 1/2 years at the 0.13 m^3/s (3 mgd) plant, and nearly 1 year (by February 1987) at the smallest plant. Sludge is removed manually with pitchforks at two of the plants. The total labor time required is 9 to 10 man-hours per bed. At the third plant, concrete strips in the bed allow the use of a front-end loader to remove sludge. Use of the front-end loader cuts the labor time to only 3 man-hours per bed.

Plant administrators have plans to eliminate the use of drying beds; rainy weather severely hampers dewatering on the sand beds and there are neighbors within about 180 m (200 yd) of the plants who occasionally complain about odors. A belt press in a trailer has recently been purchased. The belt press has dewatered the sludge to greater than 20 percent solids. It is uncertain if a belt press or centrifuge will be used for future mechanical dewatering. Dewatered sludge is currently hauled about 16 to 19 km (10 to 12 mi) to a landfill for disposal.

8.2.4 Chicago, Illinois

The Metropolitan Sanitary District of Chicago operates the West-Southwest Wastewater Treatment Plant (WSW Plant). Following is a description of the plant as it operated in 1986. The plant has an average daily design capacity of 53 m^3/s (1,200 mgd) of activated sludge secondary treatment and a design capacity of 26 m^3/s (600 mgd) of Imhoff tanks available for primary treatment. In addition, there are conventional primary clarifiers with a maximum hydraulic design capacity of 48 m^3/s (1,100 mgd). The average daily flow rate is about 35 m^3/s (800 mgd), divided nearly evenly between the Imhoff tanks and the remainder of the WSW Plant. There is an annual average Imhoff tank sludge solids production of about 64 Mg/d (70 tons/d); of this, about 36 to 41 Mg/d (40 to 45 tons/d) of solids are applied to sand drying beds, and about 23 to 27 Mg/d (25 to 30 tons/d) of solids are applied directly to lagoons.

Polymer has been used nearly continuously to aid dewatering on the sand drying beds at the WSW Plant since about 1969-1970. Sludge is applied to the drying beds regularly from April until November. For the winter period, the last sludge application is usually in December, with a top-off application in January/February. Dewatered sludge is removed with a diesel-powered digging machine that operates on a system of rails. This digging machine is also used to break up the surface crust that forms on the drying sludge. The original rail sludge removal system was installed in 1931.

A liquid cationic polymer is used to aid dewatering. The polymer is diluted with 9 l of water/l of polymer and is stored in a 360,000-l (94,000-gal) tank. A metering pump discharges polymer solution into the sludge lines as the sludge flows to the drying beds.

The digested Imhoff tank sludge is removed at an average solids content of about 5.5 to 6 percent. Sludge is applied on the sludge beds to an average depth of about 28 cm (11 in). There are 12 sand drying beds and a total bed area of 11.0 ha (27.3 acres). The average polymer dosage applied to the sludge beds is 15 to 20 g of liquid cationic polymer/kg (20 to 40 lb/ton) of solids. The polymer cost is $0.17/kg ($0.075/lb) as delivered, which results in a polymer cost of $2.50 to $3.30/Mg ($2.25 to $3.00/ton) of solids.

Dewatered sludge is removed from the sludge beds after an average non-winter drying time of 35 days during 1985. This drying time is probably several days longer than normal since 1985 was considered to be a wet year. The average dewatered sludge solids content during 1985 was 33 percent. Dewatered sludge is removed from the plant site by rail car and is stored temporarily in piles in a storage area off-site. The sludge is used by various non-agricultural users, by the Parks District, and for reclamation of completed sanitary landfills as a topsoil material.

During 1985 a total of 12,468 Mg (13,744 tons) of dry sludge solids were applied to the sand drying beds. This total results in an annual solids loading of 113 $kg/m^2/yr$ (23.1 lb/sq ft/yr). The total operation and maintenance costs for operation of the sand drying beds during 1985 are as follows:

Item	Cost
Polymer	$3/Mg ($3/ton) dry solids
Maintenance	$36/Mg ($33/ton) dry solids
Labor	$21/Mg ($19/ton) dry solids
Total O&M Cost	$60/Mg ($55/ton) dry solids

One reason the maintenance cost was so high in 1985 is that the digging machine used to remove the sludge is a very old piece of equipment that requires frequent repairs. A new digging machine, expected to be operational in September 1986, has an estimated cost of $1,700,000.

8.3 Paved Drying Beds

A post construction evaluation of a sludge handling system presently used by the City of Fort Worth, Texas, was prepared for the U.S. Environmental Protection Agency (1). This section is based on this evaluation. The system to be evaluated was a tractor-mounted auger aerator dewatering device. Although the drying beds in Fort Worth are unpaved sand beds, the paved bed concept is still applicable to this case study.

8.3.1 Post Construction Evaluation - Fort Worth, Texas

The City of Fort Worth purchased an auger aerator machine in March of 1984. The City has been using it since that time in their sludge handling facilities to increase the drying rate of sludge on the sludge beds, reduce insect problems, and to windrow dried sludge to facilitate removal from the sludge beds and loading in trucks.

The Village Creek Wastewater Treatment Plant is a 4.38-m^3/s (100-mgd) design flow activated sludge wastewater treatment plant, owned and operated by the City of Fort Worth, Texas. This wastewater treatment facility includes primary sedimentation tanks, diffused air activated sludge treatment, secondary clarification, effluent filtration, and chlorination prior to discharge to the Trinity River. Existing plant flow has averaged 3.87 m^3/s (88.4 mgd) and has varied between 3.36 m^3/s (76.8 mgd) and 4.75 m^3/s (108.4 mgd) over the period of October 1982 to July 1984. All sludge production and performance figures are based on monthly averages of plant performance for this 22-month period. The average BOD and suspended solids concentrations for influent sewage are 271 and 240 mg/l, respectively. Primary and secondary sludges are collected and handled separately. Primary sludge is collected and thickened to about 5 percent solids prior to digestion in six anaerobic digesters. Waste activated sludge is collected and thickened by centrifuges to an average of 3.9 percent solids prior to digestion in four anaerobic digesters.

Raw sludge production and feed to the digesters has averaged 116,000 kg/d (256,000 lb/d). After digestion and volatile solids reduction, the average dry solids to the sludge drying beds are 82,100 kg/d (181,000 lb/d). These solids are sent to the drying beds as liquid sludge at an average rate of 0.029 m^3/s (0.667 mgd), which reflects an average digested sludge solids concentration of approximately 3.25 percent solids. The drying beds are located approximately 1.3 km (0.8 mi) from the treatment plant.

The sludge drying beds are located on a 120-ha (296-ac) site and the total bed area including all 52 beds is approximately 93.5 ha (231 ac). At the present time, eight drying beds have been removed from service to provide a buffer zone between the sludge drying area and adjacent residential housing. This effectively removes 15.5 ha (38.3 ac), thereby leaving 78.0 ha (192.7 ac) of sludge beds available for use.

There are no detailed records of the sludge drying bed performance over the past years. The engineering staff felt that the available sludge drying bed area was adequate and probably matched the 4.38 m^3/s (100 mgd) design capacity of the treatment plant during wet years. In dry years the operators felt that they could complete two full drying cycles on each bed. There appears to be little or no excess capacity in the sludge drying beds.

During normal operation of the sludge drying beds, the City of Fort Worth had several concerns. There has been a constant and chronic odor problem associated with the sludge drying beds. The odor nuisance was compounded by recent construction of new houses adjacent to the sludge drying bed site. The land, which was previously zoned industrial or commercial, had been re-zoned to residential. As new homes are built and families move in, the impact of fly and odor problems at the drying bed site will become more critical. The Village Creek drying beds have experienced a severe psychoda fly problem especially in the spring months. A secondary, also less troublesome, fly problem has existed in the fall. In the summer and winter months, psychoda flies have not been a severe problem.

The dense fly population, the odor from the beds, and the close proximity of new homes combine to create a serious public relations problem. The operators report that at one time the psychoda fly population consisted of very small flies that could pass through a typical house screen. One can imagine the impact this had on neighboring residents.

The flies breed in the dry crust that forms on the top of the sludge as it sits in the drying bed. The auger aerator is used to break up this crust and mix it with the wet sludge, which has two beneficial effects. First by breaking up the upper layer of crust, wet sludge is exposed to the atmosphere to accelerate drying. It also mixes fly eggs and larva into the sludge and removes the dried crust, on which the flies land and deposit their eggs. The effect of the tractor on insect nuisance is discussed further in the qualitative analysis of the machine later in this section.

The device is a four-wheel tractor with a hydraulic motor-driven horizontal auger mounted in front. It was originally designed as a high-production earth-handling machine that would backfill cross-country pipeline trenches with loose excavated material that had been stockpiled alongside the pipeline trench. The engine is a 168-kW (225-hp) turbocharged after-cooled diesel. The tractor is a four-wheel drive unit and can be steered with the rear axle, the front axle, or with both axles. At the driver's option, both axles can be operated in tandem to "crab" the machine. The all-wheel drive permits the tractor to travel in wet sludge beds without getting stuck. A photograph of the tractor auger machine is shown in Figure 8-2.

The auger is protected from excessive wear by replaceable inserts. In practice, the auger is driven through wet sludge beds to break up the crust, and through drier sludge beds to turn the sludge and expose new surfaces for drying. According to one tractor operator, it takes the machine about 1 hour to completely mix a 2-ha (5-ac) wet sludge bed filled to its normal depth of 0.46 m (1.5 ft). It takes 2 to 3 hours to turn and windrow a drier sludge bed.

The auger aerator machine is currently handling a digested sludge drying bed input of over 82 Mg/d (90 tons/d) operating 8 hours/d and 5 days/week. The treatment plant staff indicated that they felt they could more effectively control insects if they had a second unit. However, the one unit brought great improvements in operations.

The costs of utilizing the auger aerator machine are calculated and shown below and in Table 8-3. Annual costs are estimated based upon 2,000 hours of machine operation annually. Fuel costs are based upon a fuel consumption of 2.6 liters (7 gal)/hr at a cost of $0.27/liter ($1.00/gal). Present worth is based upon a 10 percent interest rate and an equipment life of 10,000 hours, which is equivalent to 5 years. Labor cost is assumed to include a single operator at a cost of $12.00/hr.

The cost per metric ton of dry solids is determined by dividing the total annual cost by the mass of dry metric tons influent to the beds, annually at 29,900 Mg dry solids/yr (33,000 tons/yr). Thus, the cost per metric ton was computed to be $3.54/Mg ($3.20 per ton) over the life of the tractor.

The Village Creek Wastewater Treatment Plant has received several benefits from the use of the tractor

Figure 8-2. Auger aerator dewatering machine.

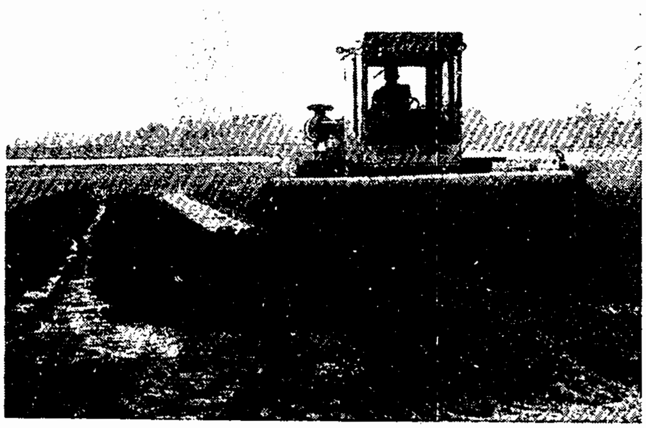


Table 8-3. Costs for Auger Aerator Tractor Used To Improve Sludge Drying Bed Operations at Village Creek Wastewater Treatment Plant

Item	Estimated Cost ($)
Capital Costs	
Auger Aerator Tractor	200,000
Annual Cost Equivalent[a]	52,800
Annual O&M Costs	
Operation	
Fuel[b]	14,000
Labor[c]	24,000
Maintenance[d]	15,000
Total O&M, $/yr	53,000
Annual Capital Recovery	52,800
Total Annual Cost[e]	105,800
Cost per dry ton of solids[f]	3.20
Total Present Worth of Auger Aerator System[g]	401,000

[a] Using Capital Recovery Factor of 0.2638, based on i = 10%, 5-year life (10,000 hr @ 2,000/yr).
[b] 7 gal/hr x $1.00/gal x 2,000 hr/yr.
[c] Man-hour rate of $12/hr x 2,000 hr/yr.
[d] 7.5% of capital cost.
[e] Annual cost equivalent + total annual O&M.
[f] Total annual cost ÷ 33,000 dry tons/yr of sludge.
[g] Present worth factor = 3.791, based on i = 10%, 5-year life.

auger equipment. Insect abatement has been an important benefit, according to both engineering and operating staff at the plant. The fact that fly eggs and larva are mixed continually into and below the surface of the sludge has dramatically cut the fly population and its impact on the surrounding housing. However, the staff feels that their single machine is barely adequate to handle the 81 + ha (200 + ac) of sludge beds and mix them with sufficient frequency to break the fly cycle. The importance of this problem will increase in the near future as additional nearby residences are constructed and families move in.

Odor control is another important consideration. The drying beds cover nearly 1.3 km^2 (1/2 square mile). A piped deodorant system has been installed surrounding the sludge bed area. The system is operated on a continuous basis, but, if there are severe odor problems, effective odor masking cannot be accomplished with a deodorant system. The tractor auger is used in several ways to control odors. By mixing and turning the sludge frequently, the drying time is accelerated, thereby cutting the total acreage of wet sludge at the site, resulting in less surface area for odor release. Other options are possible with the machinery on the site. Wet sludge and dry sludge could be mixed and windrowed for composting operations. Composting helps keep sludge aerobic and thereby reduces odor potential.

Reduction of drying time is an important consideration. Such a reduction increases the capacity of the sludge beds, and can so delay the future installation of alternative dewatering processes or enlargement of the sludge beds. Thus, this machinery has lengthened the projected life of the sludge bed drying system for Fort Worth. It has made the sludge beds more effective in drying sludge, in that less area is required to dry the existing sludge (or the sludge drying capability has increased for the existing area). It has also cut down the nuisance associated with the sludge beds, which may make them more acceptable to nearby residents.

The accelerated drying times have permitted the establishment of a buffer zone between the sludge beds in use and the new housing. The City has dedicated eight sludge beds for use as a buffer zone. This zone creates a sludge-free area about 150 m (500 ft) wide between sludge beds and nearby houses. Prior to the purchase of the auger aerator machine, all sludge beds were needed during wet years to dewater the plant's digested sludge output.

Another benefit derived by the purchase and use of the tractor auger is that it is an active, visible show of

Table 8-4. Rainfall and Air Temperature Data for Air Drying Case Studies

Case Study Location	Station Location	Mean Annual Rainfall, in/yr			Air Temperature	
		Maximum	Minimum	Mean	°C	°F
Elgin, IL	Aurora, IL	n/a	n/a	35.6[1]	9.3	48.7[1]
Belleville, IL	Belleville, IL	n/a	n/a	36.8[1]	12.9	55.3[1]
Gwinnet Co., GA	Atlanta, GA	71.4	31.8	48.6[2]	16.4[2]	61.5[2]
Chicago, IL	Chicago, IL	49.4	21.8	35.2[3]	9.4	48.9[3]
Ft. Worth, TX	Dallas/Ft. Worth, TX	50.6	18.6	32.1[2]	10.7	65.6[2]

[1] Yearly average for the period 1951-1980.
[2] Yearly average for the period 1944-1983.
[3] Yearly average for the period 1958-1983.
n/a = not available.
in/yr x 2.54 = cm/yr.

good faith to nearby residents. Such equipment serves as evidence that the treatment plant is taking action to resolve the complaints of neighbors.

An additional benefit that the engineering staff foresees is that the tractor auger will still be useable if they convert to a sludge composting operation. Already, woodchips have been stockpiled in one section of the drying bed site for use as a bulking agent in pilot studies with sludge composting. The auger aerator is directly applicable for composting operations because of its mixing and windrowing capabilities.

Another benefit of the auger aerator is that operator acceptance has been very good. The machine has an air-conditioned cab with a stereo and is pleasant to operate. Because of the comfort, the operators are eager to use the unit. The "user-friendly" cab has two major benefits. First, the operators take pride in the condition and appearance of the tractor auger. At the time of this site visit in 1984 the machine was being thoroughly cleaned at the end of each day. Second, since operators enjoy using the machine, it is more likely to be used as intended. Therefore, it is more likely that the sludge beds will be maintained in a way that accelerates the drying time and reduces insect and odor problems.

It should be noted that the tractor auger also has one disadvantage. The machine may be creating a long-term problem since the auger mechanism has a tendency to scrape and mix sand from the sand drying beds with the sludge. Since the sand is well mixed with the sludge, it is removed when the sludge is trucked from the site. In the long term this removal could cause a lowering of the sand beds and could ultimately require the replacement of sand materials to avoid exposing or damaging the underdrain system. However, since the Village Creek Plant's underdrain system is said to be installed nearly 1.8 m (6 ft) below the surface, the sand removal is not an immediate concern.

8.4 Weather Data

Table 8-4 provides annual average rainfall and air temperature data for weather station locations near the locations described in this chapter. This weather data can be used to see if differences in dewatering rates between locations are caused by differing rainfall amounts and air temperatures.

8.5 Reference

1. *Technology Evaluation of Brown Bear Tractor for Sludge Dewatering*. J.M. Montgomery Inc., EPA Contract Report 68-03-1821, U.S. Environmental Protection Agency, Water Engineering Research Laboratory, Cincinnati, OH, 1984.

Chapter 9
Case Studies: Mechanical Dewatering Systems

9.1 Introduction

Six case studies on mechanical dewatering processes are presented in this chapter. They were selected to illustrate the use of these processes as well as to relate performance and operation and maintenance experiences. Where possible, case studies were selected on the basis of availability of information, data, and side-by-side comparisons of different processes. This manual does not aim to endorse any particular manufacturer's dewatering or conditioning process. It is recognized that there are, as indicated in Appendix C, other manufacturers of the same types of equipment and conditioners. Further some of the unmentioned manufacturers may have had both better and more economical experiences than those illustrated by the case studies. The case studies include:

- Stamford Water Pollution Control Facility: Stamford, CT - Parkson Belt Filter Press
- Southeast Water Pollution Control Plant: San Francisco, CA - Comparison of Sharples and Humboldt-Wedag Centrifuges
- Calumet Sewage Treatment Works, Metropolitan Sanitary District of Greater Chicago (MSDGC): Calumet, IL - Harima Centrifuge
- West-Southwest Sewage Treatment Works, Metropolitan Sanitary District of Greater Chicago (MSDGC): Calumet, IL - Sharples Centrifuge
- Metropolitan Denver Central Plant: Denver, CO - Comparison of Various Centrifugal and Belt Press Dewatering Processes with the Original Vacuum Filtration Equipment
- Duffin Creek Water Pollution Control Plant: Toronto, Canada - Jones and Edwards Membrane Filter Press using Polymer Conditioning

9.2 Case Study: Belt Filter Presses, Stamford, CT

The City of Stamford's Water Pollution Control Facility (shown in Figure 9-1) is a 0.88-m^3/s (20-mgd) conventional, activated sludge treatment plant. Its wastewater is about 15 percent industrial and 85 percent commercial/domestic. The average influent BOD_5 and suspended solids are 160 mg/l and 120 mg/l respectively. Effluent BOD_5 and suspended solids average 7 mg/l and 20 mg/l respectively.

The plant has primary sedimentation followed by mechanical aeration. The average MLSS is 2,500 mg/l, SVI is 100 ml/g, and SRT is 6 days. Sludges from the underflow of the primary and secondary clarifiers are combined, degritted using hydrocyclones, and gravity thickened. The typical underflow concentration from the thickeners is 2.5-3.0 percent solids.

Stamford is unique in that it has the only operational co-incineration system for the disposal of sludge and municipal refuse in the United States (see Figure 9-2). Using progressive cavity pumps, the sludge is pumped to flocculation tanks. Calgon WT-2136 polymer, which is used to condition the sludge, is added approximately 7.6 m (25 ft) upstream of these tanks. The conditioned sludge is dewatered and the cake is then discharged to the pug mill. There the dewatered sludge is combined with previously dried sludge to produce a mixture which is approximately 60-65 percent solids. This mixture is discharged from the pug mill and conveyed to the rotary dryer. A portion of the hot gas which would normally be wasted through the incinerator stack enters the dryer at the same area as the sludge mixture. As the dryer rotates (5 rpm), the sludge is cascaded through these hot gases, thus evaporating the moisture in the sludge. The dried sludge (90 percent solids) is discharged through a diverter gate and divided into two streams - one that goes back to the system as dry recycle and the other that goes to the incinerator and is burned. The heat value of the sludge is about 12,800 kJ/kg (5,500 Btu/lb).

When the plant was built, centrifuges were installed for dewatering. However, they were unable to achieve the requisite 22 percent solids needed to allow optimization of the sludge drying and incineration system. Also, they were extremely high maintenance items and the solids capture with polymer addition was less than 60 percent. In 1979, these units were

Figure 9-1. Schematic of sewage treatment plant, Stamford, CT.

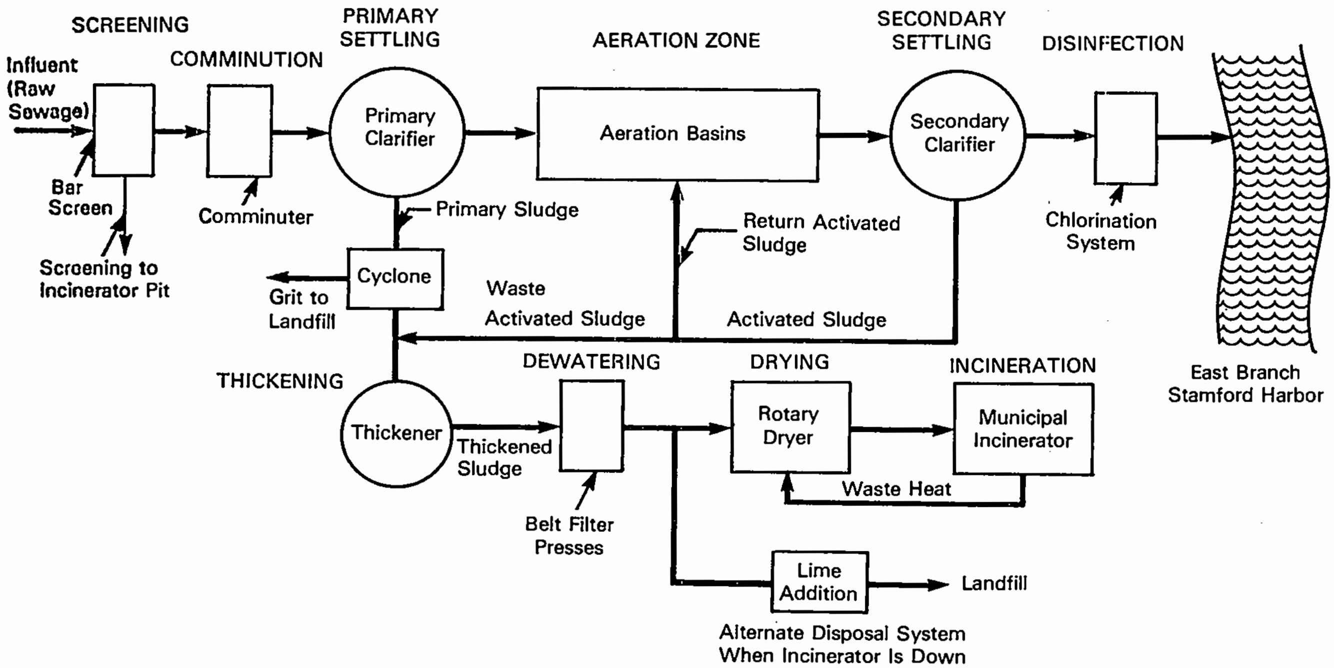
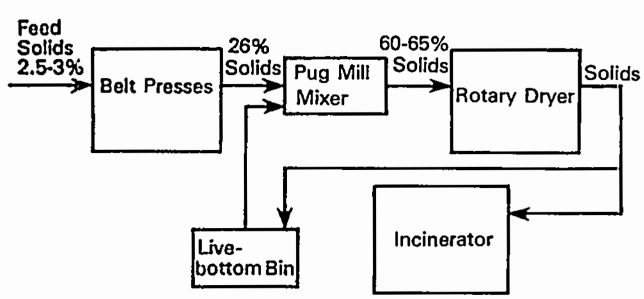


Figure 9-2. Sludge flow path through co-incineration system.



replaced with two 2-meter Parkson MP-80 belt filter presses. These units were equipped with an additional high pressure section to ensure the driest possible sludge cake. The City specified a minimum cake solids concentration of 24 percent with a 95 percent capture.

During the first two months of operation, cake solids concentrations of less than 20 percent were produced but the solids capture was an excellent 98 percent. In the third month of operation, the cake solids concentration began to increase and rose by the end of the month to 25-26 percent. Solids capture remained at 98 percent. It was theorized that, because the centrifuges had such an extremely poor solids capture, fine solids were continuously circulating through the plant. With the high solids capture efficiency of the belt filter presses, these fines were gradually being removed. Once that occurred, the sludge dewatered more easily and a drier cake was produced.

Initially, belt or screen life was poor (less than 500 hours). The short screen life was attributed to operator inexperience and a sludge that was somewhat difficult to dewater during that period. Experimentation with different mesh screens led to the selection of the Scandiafelt FE-3366 screen.

This screen has a high air permeability and allows for good drainage of water from the sludge. This feature prevented uneven buildup of the sludge prior to the extra-high pressure section, thus preventing creasing and wearing of the screen cloth.

Typical maintenance problems for these units were failure of ballbearings and occasionally broken roller shafts. Bearing life increased substantially when the press operators were given the responsibility of greasing the bearings.

In 1985, the MP-80 units were traded in on the improved Parkson Series 3000 presses. They have been operated for 24 hours per day on an average of 6 days per week for over 2 years (15,000 hours) with no mechanical failures at all (bearings, shafts, drive units). Screen life has been outstanding with upper screen lives of as much as 10,000 hours and lower screen lives of about 5,000 hours. At one time, a series of lower screens failed at the clipper seams after only 200-300 hours of operation. However, this failure was due to a problem with the belt manufacture and not with either the presses or their operation.

Currently, with a raw feed sludge solids of 2.5-3.0 percent (50:50, P:WAS), a cake solid of 27 percent is achieved with 98 percent capture. Approximately 40 kg of Calgon WT-2136 cationic polymer per dry Mg (80lb/dry ton) of sludge is used at a cost of \$13.22/dry Mg (\$12.00/dry ton). In addition, potassium permanganate is added to the feed sludge at the rate of 0.5 kg/Mg (1 lb/dry ton) for odor control. This adds about \$1.10/dry Mg (\$1.00/dry ton) of sludge, but the cost is justified by giving the operators a better working environment.

9.3 Case Study: Centrifuges, San Francisco, CA

9.3.1 Sludge Characteristics and Processing

The San Francisco Southeast Water Pollution Control Plant (WPCP) is a 3.5-m^3/s (81-mgd) pure oxygen activated sludge wastewater treatment plant. Pertinent design features include:

Primary clarifiers:
7 @ 12 m x 64 m x 3.7 m SWD
(38 ft x 210 ft x 12 ft SWD)

4 @ 11 m x 78 m x 3.4 m SWD
(37 ft x 256 ft x 11 ft SWD)

Aeration basins:
8 @ 1.8-13 m x 13 m x 4.3 m SWD
(6-42.5 ft x 42.5 ft x 14 ft SWD)

Secondary clarifiers:
16 @ 36.6 m dia. x 4.6 m SWD
(120 ft dia. x 15 ft SWD)

The 1986 operational characteristics of the plant are as follows:

	Influent	Primary Effluent	Secondary Effluent
Flow, mgd	81	--	- -
BOD_5, mg/l	201	127	11
TSS, mg/l	246	106	18
TP, mg/l	--	--	3
NH_4N, mg/l	--	--	21
Temp.	21°C	--	- -

The operational characteristics of the activated sludge include a HRT of 1.8 hours and an estimated 1.9 day mean cell residence time (MCRT) and 0.7 day^{-1} F/M. The pure oxygen activated sludge system is operating in a non-nitrifying mode of operation. The aeration system consists of six stages of complete mix reactors in each of the eight trains. A schematic for the San Francisco Southeast WPCP is shown in Figure 9-3.

The excess primary and biological sludge produced from the plant is estimated to be as follows:

PS -	43,090 kg/d (95,000 lb/d)	60%
WAS -	29,485 kg/d (65,000 lb/d)	40%
TOTAL -	72,575 kg/d (160,000 lb/d)	100%

The primary clarification efficiency is 55 percent removal of TSS and 35 percent removal of BOD_5. The high efficiency of primary treatment results in a 60:40 ratio of PS:WAS in the sludge mixture. The sludge yield is about 0.70 lb EAS/lb BOD_5 removed based on 5 mg/l $SBOD_5$ in the final effluent. The SVI averages about 150 ml/g and normally ranges from 100 to 200 ml/g. Sludge is anaerobically digested prior to centrifugation.

9.3.2 Centrifuges

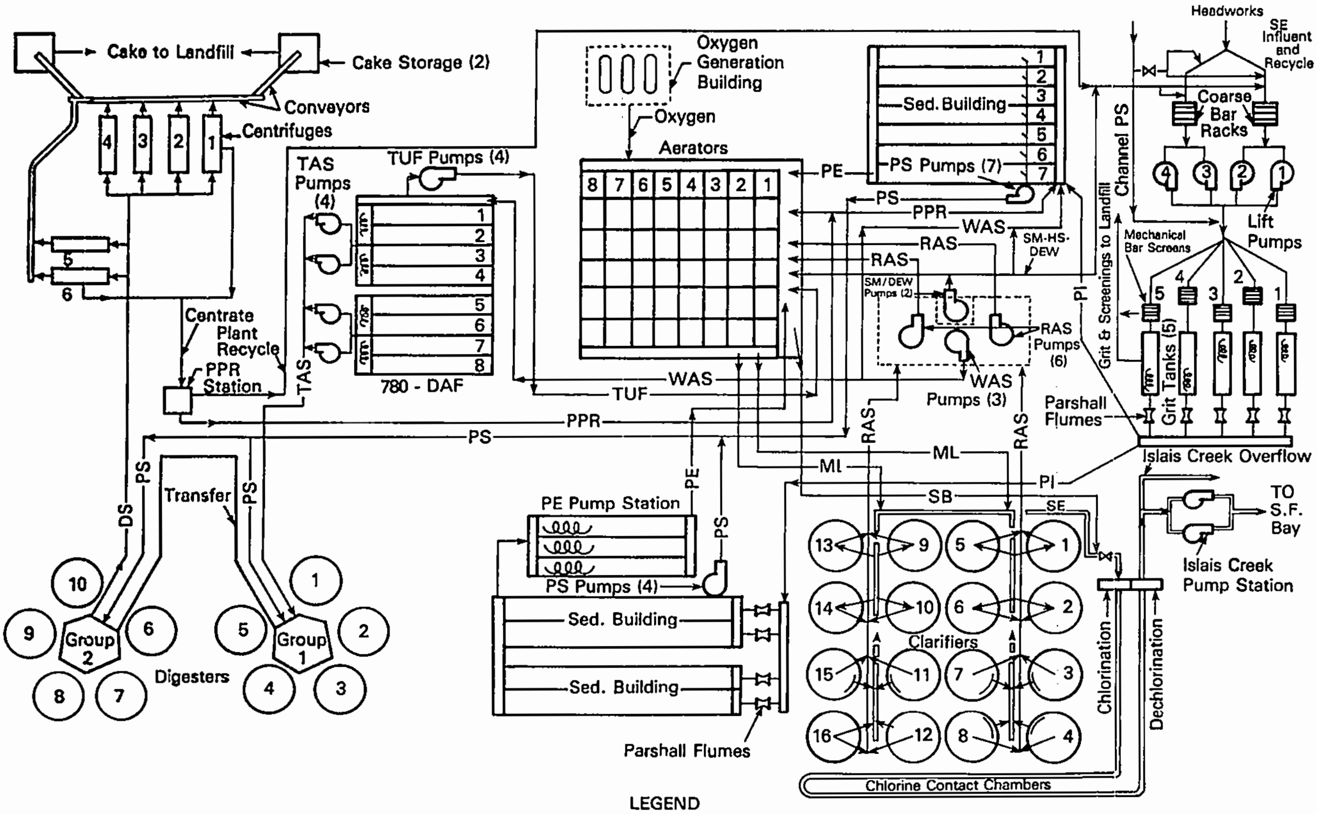
Two Sharples PM75000 centrifuges were installed in 1982 and two more were added in 1985. These countercurrent flow centrifuges are driven by 112-kW (150-hp) motors and are equipped with manually controlled eddy current backdrives. The PM75000 has a 0.7-m (29-in) bowl diameter and a 2.3-m (92-in) bowl length and operates at 2,300 rpm and 2,500 rpm or a peripheral force of 2,180 to 2,500 g's. The differential speed is normally in the range of 2.5-4.0 rpm. The bowl and scroll are fabricated from ACI CFB SS and the housing is 316 SS with a cast iron base. The scroll tips and other wear points are protected by replaceable tungsten carbide tiles.

Sludge is fed to the centrifuges by 0-15.8 l/s (0-250 gpm) variable flow progressive cavity pumps. The feed rate is normally 9.46 l/s (150 gpm) and is adjusted to account for varying sludge concentrations and characteristics. Polymer is added to the centrifuge in the bowl just downstream of the feed port. Cake is removed from the centrifuge discharge hopper with progressive cavity sludge cake pumps and a cake conveyor.

The Humboldt-Wedag centrifuge installation consists of two Model S4-1 concurrent flow units which started up in December 1982. The S4-1 centrifuges have a bowl diameter of 0.9 m (36 in) and a bowl length of 2.4 m (96 in). The bowl speed is 1,400 rpm (1,000 g's) and the units are equipped with automatic torque controlled hydraulic backdrives. The differential speed is normally in the range of 2 to 4 rpm. The bowl and scroll are fabricated from carbon steel. The housing and base are made of carbon steel. The scroll and other wear points have Udalite and/or ceramic hardfacing.

Sludge feed to the centrifuges is controlled by two 0-15.8 l/s (0-250 gpm) sludge pumps. Normal feed

Figure 9-3. Schematic of the Southeast WPCP, San Francisco, CA.



LEGEND

PPR = Pumped Plant Recycle
RAS = Return Activated Sludge
TAS = Thickened Activated Sludge
TUF = Thickener Underflow
WAS = Waste Activated Sludge
DS = Digested Sludge
ML = Mixed Liquor
PE = Primary Effluent
PI = Primary Influent
HS = Heavy Solids
SM/DEW = Scum/Dewatering
PS = Primary Sludge
SB = Secondary Bypass
SE = Secondary Effluent
SI = Secondary Influent

rate is 9.46 l/s (150 gpm). Polymer is added to the feed zone area of the centrifuge. Centrifuge cake is transferred from the centrifuge hopper with progressive cavity pumps. Ferric chloride is added to the sludge upstream of all centrifuges for struvite (magnesium ammonium phosphate scale) control.

The centrifuges were evaluated by City personnel and technicians and engineers from the centrifuge suppliers. Centrifuge setup was optimized by the suppliers prior to the test conducted in mid-1984. The performance test results are presented in Tables 9-1 and 9-2. The tests employed Centrifuge No. 1 (Sharples) and Centrifuges 2 & 3 (Humboldt-Wedag) for tests conducted on the same feed of PS and WAS. The ratio of PS:WAS was about 1:1 at the time of the tests. (Note that these tests are not necessarily representative of current performance and that one of the requirements of the test was a minimum 95 percent recovery).

In Figure 9-4, the respective cake solids as a function of the scroll shaft torque is presented. As indicated, the driest cake is produced at the highest torque, which generally represents the maximum residence time of solids in the bowl. This conclusion is somewhat modified by other factors such as feed rate, solids discharge rate, solids recovery, and polymer dosage.

Both types of centrifuge demonstrated a decline in cake solids with increasing feed rate. Since the feed TSS was relatively constant, the probable effect was increased solids discharge rate as a function of feed rate. Both effects on cake solids are shown in Figure 9-5. In this case, if the design objective was to maximize the cake solids, it would be necessary to operate the centrifuge at less than the maximum capacity, possibly to 50-70 percent of maximum capacity.

Table 9-1. 1984 Centrifuge Performance Test, Sharples Centrifuge No. 1 - San Francisco Southeast WPCP*

	6/4			6/5		6/6	6/7			6/8	
Feed Flow, gpm	150	150	148	194	199	250	125	126	126	173	173
Feed, %TS	1.9	1.9	1.9	1.9	1.9	1.9	1.8	1.9	1.8	1.8	1.8
Feed, %VS	67.3	67.7	67.9	69.4	69.7	67.7	68.4	68.0	70.0	70.4	68.6
Polymer Conc., %TS	0.30	0.39	0.32	0.34	0.34	0.22	0.21	0.21	0.17	0.23	0.24
Polymer feed, gpm	3.8	4.6	6.4	5.0	6.5	8.0	2.4	4.0	5.0	6.0	7.5
Polymer dose, lb/ton	7.9	12.4	14.4	8.9	11.4	7.4	4.5	7.0	7.5	8.8	11.3
Cake %TS	18.4	19.7	22.2	16.5	18.9	13.9	14.4	17.2	19.5	19.2	21.7
Centrate TSS, mg/l	540	533	860	1,100	1,270	1,060	1,570	1,120	920	660	1,140
Recovery, %	97.5	97.5	95.8	95.0	94.1	95.2	92.3	94.7	95.3	96.7	94.3
Energy, kWh	60.8	58.1	-	68.9	72.6	48.2	52.4	53.8	54.1	67.0	68.4
Energy, kWh/gpm	0.41	0.39	-	0.36	0.36	0.3	0.42	0.43	0.43	0.39	0.4
Torque, kg/m^2	333	369	486	360	440	312	280	330	350	390	450

* Bowl speed: 2,300 rpm
Polymer: Percol 757
Pond depth: not reported.

Table 9-2. 1984 Centrifuge Performance Test, Humboldt Centrifuges Nos. 2 and 3 - San Francisco Southeast WPCP*

	6/5 #2		6/6 #2	6/7 #3			6/8 #3	
Feed Flow, gpm	188	193	247	125	126	125	171	172
Feed, %TS	1.9	1.9	1.9	1.8	1.9	1.8	1.8	1.8
Feed, %VS	69.4	69.7	67.7	68.4	68.0	70.0	70.4	68.6
Polymer Conc., %TS	0.34	0.34	0.22	0.21	0.21	0.17	0.23	0.24
Polymer feed, gpm	5.5	6.3	8.0	2.4	4.0	5.0	6.0	7.5
Polymer dose, lb/ton	10.1	11.4	7.5	4.5	7.0	7.6	8.9	11.4
Cake %TS	17.8	17.3	14.7	11.9	19.0	22.8	21.7	20.9
Centrate TSS, mg/l	2,620	2,650	2,220	1,850	1,320	980	580	1,700
Recovery, %	87.8	87.6	89.7	91.1	93.7	95.0	97.1	91.5
Energy, kWh	41.3	41.9	48.2	31.9	32.2	32.1	37.0	37.4
Energy, kWh/gpm	0.22	0.22	0.19	0.26	0.26	0.26	0.22	0.22
Torque, ib-in	50	60	30	20	50	80	80	90

* Bowl speed: 1,400 rpm
Polymer: Percol 757
Pond depth: 520 mm.

Both types of centrifuge exhibited sensitivity of cake solids concentration to the polymer dosage. While the two lowest cake solids results were due to underdosing, there was a significant improvement in cake solids with an increased dosage. As shown in Figure 9-6, the increased polymer dosage of 50 percent beyond the amount necessary to achieve 85-90 percent recovery can result in a 3-5 percentage points increase in cake solids. This may be only true for a specific polymer, in this case Percol 757. The lower cake solids again generally resulted from the highest solids rate.

At a polymer dosage of 14-18 kg/dry Mg (7-9 lb/dry ton) of solids, the Humboldt centrifuge produced 1-3 percentage points drier cake solids at equivalent feed rates. While the Sharples cake concentration increased with dosages up to 22-24 kg/Mg (11-12 lb/ton), the Humboldt cake became slightly wetter with dosages exceeding 18 kg/Mg (9 lb/ton). For these tests Humboldt modified the polymer injection with a nozzle that created a 5.6 kg/cm^2 (80 psi) pump discharge pressure. San Francisco no longer runs the system in this manner.

In general the Sharples unit had a slightly better recovery than the Humboldt centrifuge at the equivalent polymer dosage. The lowest recovery measured was 87.6 percent for the Humboldt centrifuge and 92.3 percent for the Sharples unit. Average recovery was 91.7 percent and 95.3 percent for Humboldt and Sharples, respectively. While the

Figure 9-4. **Cake solids vs. torque.**

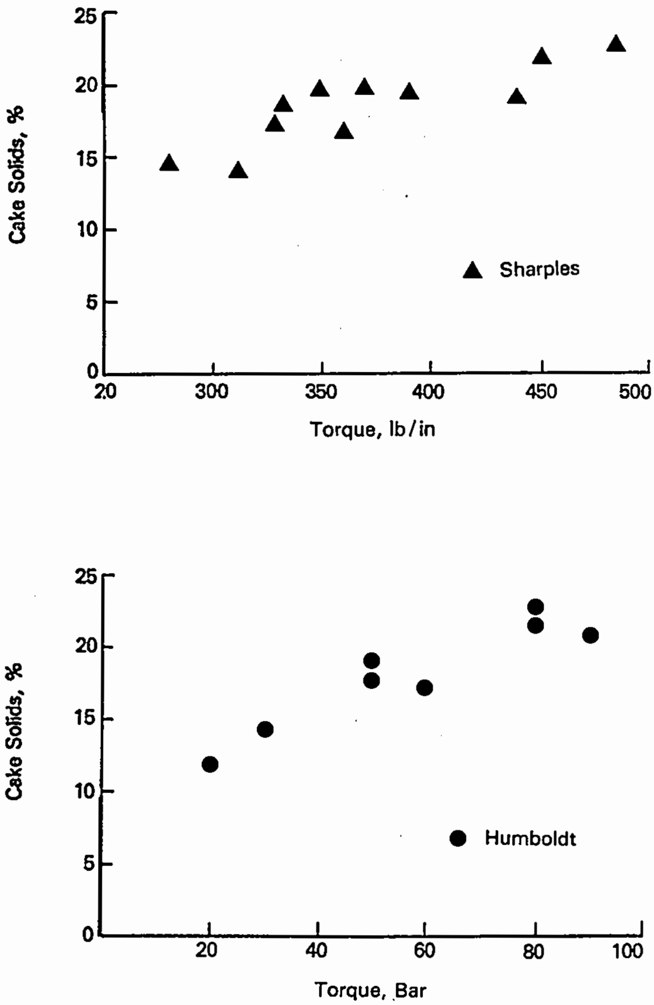


Figure 9-5. **Cake solids vs. feed rate.**

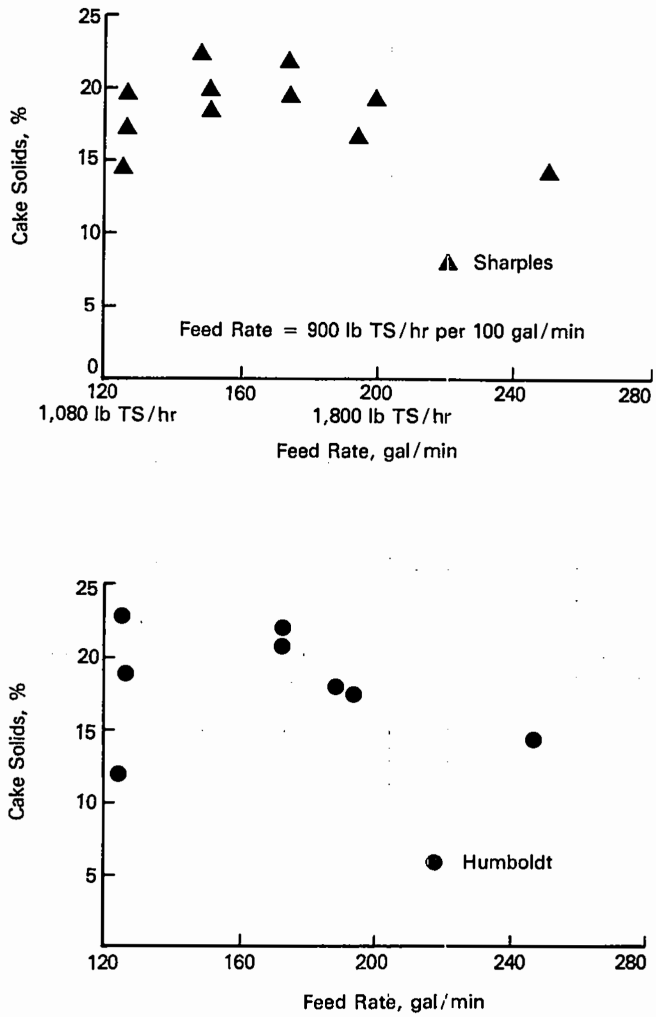
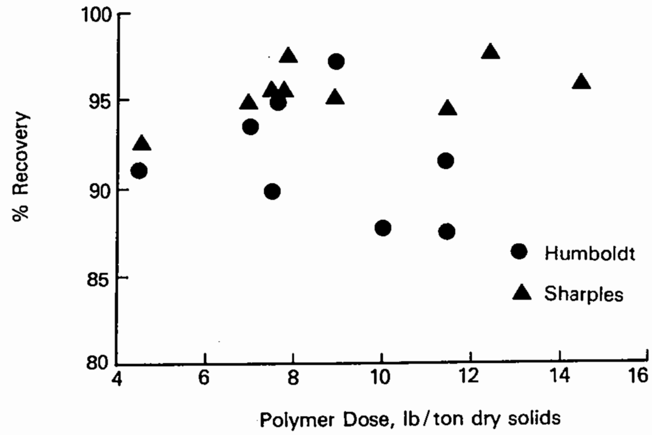
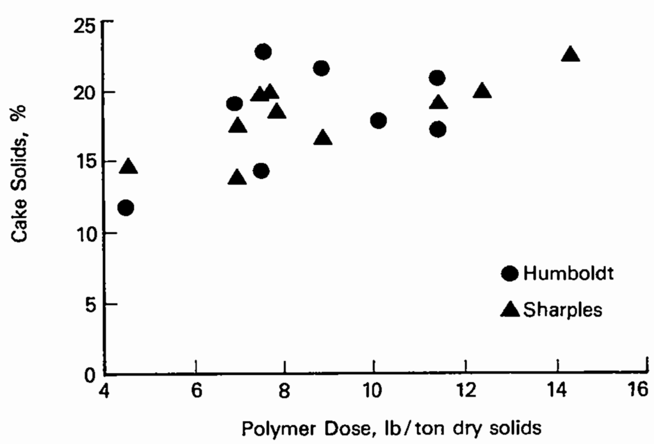


Figure 9-6. Effect of polymer dose on cake solids.



recovery data for Sharples indicated that the polymer dosage could be reduced, the reduction in recovery would be at the expense of a wetter cake as shown in Table 9-1. The recovery versus polymer dosage data is presented in Figure 9-7.

Figure 9-7. Effect of polymer dose on solids recovery.



9.3.3 Operation and Maintenance

The operating hours on the centrifuges as of September 1986 are as follows:

Centrifuge No.	Manufacturer	Date Installed	Operating Hours
1	Sharples	1982	13,368
2	Humboldt	1982	17,415
3	Humboldt	1982	20,711
4	Sharples	1982	14,903
5	Sharples	1985	3,912
6	Sharples	1985	1,638

The plant reports that there has been no appreciable wear on the scrolls of either machine. Most of the wear has occurred in the area of feed inlet and sludge cake discharge. A detailed breakdown of costs is not available. The total amount spent up to September 1986 is about $450,000 for labor and materials - $300,000 for Sharples and $150,000 for Humboldt. However, the $300,000 for Sharples includes the cost to refurbish machines purchased from New York City. The gross maintenance cost of the centrifuges (including refurbishing of New York City machines) is $6.25/hr of operation. At 11.0 l/s (175 gpm) and 2 percent TSS, the cost would be $7.86/Mg dry solids ($7.13/ton).

The operating costs for power for the centrifuge are as follows:

Flow	Sharples			Humboldt		
(gpm)	kWh/gpm	$/hr	$/ton	kWh/gpm	$/hr	$/ton
125	0.43	3.76	6.68	0.26	2.28	4.05
175	0.40	4.90	6.22	0.26	2.70	4.33
250	0.31	5.43	4.82	0.20	3.50	3.11

Based on $0.075/kWh, sludge @ 2% TSS, and 85% recovery. The Sharples recovery is estimated to be 90% and the Humboldt recovery is estimated to be 80%.

The chemical dosage and costs for the centrifuge operation are as follows:

Polymer (Percol 757 - Allied Colloids):
8.9 lb/ton @ $2.25/lb = $20.03/ton

$FeCl_3$:
60 lb/ton @ $0.12/lb = $6.00/ton

Total = $26.03/ton ($28.69/Mg)

There are two operators per shift responsible for the centrifuge operation. Their duties include operation and adjustment of the centrifuges and auxiliary systems, batch polymer makeup, monitoring the sludge hopper level, and operating the cake conveyors, as well as some housekeeping.

The operating staff's comments regarding their experiences with the two types of centrifuges, high speed countercurrent and low speed concurrent, are summarized in Table 9-3.

Currently, the percent excitation (Sharples) needs to be monitored as it relates to the inventory of cake in the bowl and requires a manual adjustment of the backdrive. However, San Francisco intends to start up a programmable controller supplied by Sharples to monitor the percent excitation and adjust the backdrive automatically.

Table 9-3. **Operator Comments on Humboldt and Sharples Centrifuges - San Francisco Southeast WPCP**

Centrifuge Type	Advantages	Disadvantages
Humboldt	• Lower initial cost • Less power required • Simpler design, easy to operate, fewer hours for operators • Automatic torque control backdrive accomodates varying feed solids • Larger tolerances • More operating hours between servicing	• Recovery is not as high with identical polymer dosage • Hydraulic system subject to hydraulic fluid leakage
Sharples (automatic PC backdrive control is being installed)	• Wear items can be replaced on site • Will use less polymer when operating properly • Service oriented, local representatives • More gauges and monitoring devices available	• More sensitive to operational upsets, e.g., plugged cake pumps backing into machine • Needs more operator attention

At San Francisco, the higher speed machine has shown greater wear due to the speed and softer stainless steel, and also due to San Francisco's fine grit (100-200 mesh) in the sludge slurry. However, this problem is partly offset by the higher speed machine's ability to capture more of a high volatile sludge introduced by the activated sludge.

9.4 Case Study, Centrifuges, Calumet, IL, Calumet STW

The Metropolitan Sanitary District of Greater Chicago (MSDGC) operates seven sewage treatment works, of which four have complete solids processing systems. The two largest of these are the Calumet Sewage Treatment Works (STW) and the West-Southwest STW. The Northside plant purges sludge to the West-Southwest plant. The dewatering operations at the two largest of the MSDGC's sludge treatment works are described below.

9.4.1 Sludge Characteristics and Processing

The Calumet STW is a 9.6-m3/s (220-mgd) single-stage conventional activated sludge plant. Pertinent design features include

Primary clarifiers:

32, each @ 1,020 m^3 (270,000 gal)
total surface area - 9,510 m^2 (102,400 sq ft)

Aeration basins:

31, total volume - 192,800 m^3 (51.1 Mgal)

Final clarifiers:

32, total volume - 179,600 m^3 (47.6 Mgal)
total surface area - 41,090 m^2 (442,300 sq ft)

The operational characteristics of the plant, based on 1985 yearly averages, are as follows:

	Influent	Primary Effluent	Secondary Effluent
Flow, mgd	235	--	- -
BOD_5, mg/l	151	131	19
TSS, mg/l	303	215	21
NH_4N, mg/l	14.7	--	12.4
Temp., °C	14	--	- -

Based on 1985 figures, the plant averages an SRT of 8.6 days and an F/M ratio of 0.14 lb BOD_5/lb MLSS/d. Since 1982, the plant has installed additional primary clarifiers, aerated grit chambers, new fine screens, and a new secondary system.

The plant anaerobically digests the primary and waste activated sludge in the following percentages:

PS -	51,710 kg/d (114,000 lb/d)	33.7%
WAS -	101,600 kg/d (224,000 lb/d)	66.3%
TOTAL -	153,310 kg/d (338,000 lb/d)	100.0%

The primary clarification efficiency is 13 percent removal of BOD5 and 29 percent of the TSS. The SVI averages approximately 110 ml/g.

9.4.2 Centrifuges

A total of five Ishikawajima-Harima Heavy Industries (IHI) centrifuges, model number HS-805M, were installed in 1981 at the Calumet plant. The units are driven by 112-kW (150-hp) motors and are equipped with manually-controlled backdrives. The bowl speed is 1,560 rpm (1,750 g's) and the normal differential speed is 12 rpm. The inside bowl dimensions measure 800 mm (31.5 in) in diameter and 2,650 mm (104 in) in length.

The units have a rated capacity of 33.2 l/s (525 gpm). The digested sludge is fed to the centrifuges by three 7.5-kW (10-hp) feed sludge pumps at a rate of 11.4 l/s (180 gpm) and an average feed concentration of 2.7 percent of total solids. Polymer is injected into the sludge at a rate of 0.2 l/s (2 gpm) just before the feed enters the centrifuge. The average polymer dosage is 5 kg dry solids/Mg (9.9 lb/ton). However, the plant notes that the centrifuges require a significantly higher polymer dosage when the concentration of the feed sludge drops below 2.4 percent of total solids.

At a 2.7 percent feed concentration and a polymer dosage of 5 kg dry solids/Mg (9.9 lb/ton), the units produced the following results:

	Cake Solids, percent			Solids Recovery, percent		
	min.	max.	avg.	min.	max.	avg.
Summer	14.5	16.7	15.6	91.0	95.4	93.8
Winter	15.2	16.2	15.7	-	-	-

September 1986 operating data for each of the five centrifuges are presented in Table 9-4.

Cake is transferred from the centrifuge discharge hopper by conveyor belts and removed from the plant by trucks. Baffles were added to the chutes on the centrifuge discharge hopper to prevent the cake from splashing off the conveyor belts. The trucks have also been modified to reduce leakage of wet cake, but trucks with tailgates prove less efficient. Pumping or screw conveyor systems are currently being considered for centrifuge cake transfer. Pumping appears to be the cleanest operation and several systems were successfully tested in 1986. No decision regarding the installation of a pump system has been reached at this time. The plant also installed secondary belt scrapers at the discharge end of each conveyor to improve the capture of solids. Previously, the solids ended up in the building drain system.

9.4.3 Operation and Maintenance

Each of the five centrifuges at the Calumet complex are operated between 3,500 and 4,200 hours per year. The IHI scrolls had to be retrofitted with sintered tungsten carbide tiles to extend the life to the 10,000-12,000 hr range. Currently, the scrolls on at least one unit are repaired each year at a cost of $75,000 for sintered carbide tiles. While the plant expects future repair costs of the scrolls to decrease, an additional $15,000 is generally estimated for in-house support costs. Over 60 percent of these costs were associated with retrofitting the tiles.

At least one set of bearings is replaced each year, normally after 10,000 hours of operation or whenever the scroll is removed for maintenance. The cost of five main bearing sets is $37,500 and in-house support costs can be as high as $8,000.

The 1985 sludge dewatering costs were estimated to be as follows:

Sludge Dewatering	$/ton Dry Solids
Polymer (Percol 763): @ 9.9 lb/ton dry solids =	15.84
Power @ 123 kWh/ton dry solids + Water @ 3,370 gpm/hr + Maintenance =	35.00
Labor =	22.00
Total =	72.84

The total cost, including transport and placing of the cake in the landfill, ranges from $83 to $110/Mg dry solids ($75-100/ton).

The Calumet plant has made several modifications to improve the operation and maintenance of the dewatering complex. Some of these changes include:

Drainage System: Since it was not feasible to replace internal building drains, which later proved to be too small, the clean water sump pumps were replaced with solids-handling trash pumps. Also, a new line replaced the main drain pipeline, which recycled drainage to the treatment plant, to prevent further settling and clogging. Areas in and around the centrifuge complex were paved to prevent clogging of street drains. The pavement also makes it easier to clean up sludge spills.

Polymer Transfer and Mixing Systems: To reduce losses and allow quicker transfer of material, the 5-cm (2-in) liquid polymer transfer piping was replaced with 15-cn (6-in) piping. The "star feeder" for metering the dry polymer into the transfer system frequently jammed and was modified. The plant also modified the polymer wetting and mixing system to ensure proper initial wetting of the material. Sufficient water quantity and pressure, coupled with adequate clearance at all points, prevent partially wetted polymer from gumming up the small openings into the feed system. The plant also converted the polymer mixing liquid from treatment plant effluent to city water. This change resulted in reduced polymer usage and less frequent maintenance of the sight glasses.

Digested Sludge Feed System: The feed surge channel was raised 1.2 m (4 ft) and, as a result, the enlarged channel now provides better suction to the feed pumps. The plant also installed larger pumps, equipped with shear impellers to minimize clogging, to handle the maximum sludge feed rate.

Several other areas still require additional modifications to optimize the operations at the Calumet plant. Two of these concerns are noted below.

Struvite Formation: Although the conversion from plant effluent to city water significantly reduced the rate of struvite formation, this crystalline product (magnesium ammonium phosphate) continues to cause problems at the centrifuge complex. The plant has made other modifications in its operations, but their effect on reducing the rate of struvite formation has been negligible. These changes included:

- Using a polymer in solution with a pH <7.0.
- Injecting a chelating agent into the feed sludge.
- Flushing water (for dilution purposes) in the centrate lines.

Table 9-4. Calumet STW: Sample of Operating Data (September 1986)

Centrifuge	Average Sludge Feed		Polymer Dosage	Solids Recovery	Cake Solids
	mgd	percent solids	lb/dry ton	percent	percent
No. 1	0.255	2.54	10.2	95.2	14.9
No. 2	0.260	2.54	8.4	95.9	16.1
No. 3	0.220	2.54	9.4	94.3	14.6
No. 4	0.230	2.54	12.3	92.9	16.4
No. 5	0.256	2.54	9.4	95.6	15.1

Plant personnel remarked that the addition of anti-scalants to the polymer proved unsuccessful because the anti-scalants deactivated the basic polymer, to the point where it was ineffective as a dewatering aid. The ultimate solution appears to entail the removal of soluble phosphorus by iron (or other heavy metal) precipitation. This additional step would decrease the pH of the centrate, which, in turn, would reduce the rate of struvite formation.

Hydrogen Sulfide Gas Formation: The Calumet centrifuge complex experienced two major hydrogen sulfide gas incidents in 1986. This toxic gas was stripped from the sludge during the centrifugation process in excessive concentrations. Although the ultimate solution lies within the treatment plant (or at the generation point of the dissolved sulfides into the sewer system), the control of the gas at both the digester and centrifuge complex is very important. Preliminary testing indicated that the levels of hydrogen sulfide gas can best be controlled by the addition of zinc chloride or ferric chloride into the sludge stream at either the digester or centrifuge complex. However, some have suggested that power venting of the centrate and cake zones would be the best solution.

Once the majority of the deficiencies were corrected, the Calumet centrifuges have operated fairly reliably. They are capable of producing a good sludge cake and an excellent centrate with reasonable polymer requirements, provided that the feed sludge solids concentration remains greater than 2.5 percent. The centrifuges themselves have proven to be good machines.

However, the problem at the Calumet complex is that overall cost of centrifugation is high. Contributing to this high cost are machine malfunctions, the large number of electrically- and mechanically-driven support systems, and high seasonal labor rates.

9.5 Case Study, Centrifuges, Calumet, IL, West-Southwest STP

9.5.1 Sludge Characteristics and Processing

The West-Southwest Sewage Treatment Works (STP) in Calumet, IL is a 52.6-m3/s (1,200-mgd) single-stage activated sludge plant. Peak flow averages 63.1 m^3/d (1,440 mgd). Pertinent design features include:

Primary clarifiers:
29 @ 30.5 m x 308 m x 3.4 m SWD
(100 ft x 1,011 ft x 11 ft)

108 Imhoff tanks @ 24 m x 24 m x 10-11 m SWD
(80 ft x 80 ft x 33-36 ft)

Aeration basins:
32 @ 132 m x 10 m x 4.6 m SWD
(434 ft x 34 ft x 15 ft) each pass (4 passes/tank)

Secondary clarifiers:
96 @ 38.4 m dia. x 4.3 m
(126 ft dia. x 14 ft) SWD

The operational characteristics of the plant, based on 1986 data, are as follows:

	Influent		Primary Effluent		Sec. Effluent
	West Side	Southwest Side	West Side	Southwest Side	
Flow, mgd	327	486	-	-	788
BOD_5, mg/l	153	273	105	401	9.3
TSS, mg/l	362	708	222	234	12
NH_4N, mg/l	6.2	10.1	8.0	11.0	3.5
Temp. °C	-	-	-	-	16

Based on 1986 figures, the plant averages an SRT of 5.6 days and an F/M ratio of 0.272 lb BOD_5/lb MLSS/d. The source, quantity and disposition of the raw sludges produced by the West-Southwest STP are as follows:

Type	Raw Sludge	Disposition	Location
	dry tons/d		
PS	635	Anaerobic Digestion	West Side Imhoff Tanks (Unheated)
	165	Anaerobic Digestion	West Southwest Imhoff Tanks (Unheated)
WAS	500	Anaerobic Digestion	West Southwest Digesters (Heated)
	135	Recycle	Return to Plant

The primary clarification efficiency is 66.9 percent and 38.7 percent respectively for the Southwest and West sides of the plant. The SVI averages 67 ml/g.

The estimated sludge mixture for dewatering is as follows:

PS -	32,66 kg/d (72,000 lb/d)	21%
WAS -	123,380 kg/d (272,000 lb/d)	79%
TOTAL -	156,040 kg/d (334,000 lb/d)	100%

9.5.2 Centrifuges

Eleven Pennwalt Sharples Super D-Canter PC 81000 centrifuges were first used in 1981 and an additional unit was installed in 1984. They are countercurrent flow centrifuges driven by 93.4 kW (125 hp) motors. The 'knits are equipped with changeable backdrive pulleys which allow backdrive speed differentials of 22, 16, 9, and 6 rpm. The backdrives have 11-kW (15-hp) motors, and the gearbox ratio is 98:1. One DC variable speed backdrive has been purchased and will be installed in 1987. The PC 81000 has a 0.6 m (25 in) by 2.92 m (115 in) bowl which operates at 2,160 rpm.

The bowl and scroll are fabricated from stainless steel. The base is cast iron and the cover is plexiglas. The scroll flighting (feed zone area) is protected with replaceable tungsten carbide tiles. The liquid area scroll flighting is protected with Stellite hard surfacing. Other areas susceptible to accelerated wear are protected by either replaceable tungsten carbide inserts or hard surfacing.

Feed sludge is provided by three constant feed centrifugal pumps. These pumps feed the distribution system with individually controlled feed valves at each centrifuge. These valves allow infinite flow adjustment. The feed rate is normally in the range of 9.5-11.4 l/s (150-180 gpm). Polymer is added to the centrifuge feed inside the feed zone. Each centrifuge has a 0-0.6 l/s (0-10 gpm) progressive cavity polymer pump. However, the capacity of these pumps is being increased to 0-1.3 l/s (0-20 gpm).

At a 3.51 percent TSS feed concentration and a polymer dosage of 115 kg/Mg (230 lb/ton), the units produced the following 1986 results:

	Cake Solids, percent			Solids Recovery, percent		
	min.	max.	avg.	min.	max.	avg.
Summer	11.9	17.9	15.4	70.6	94.6	88.4
Winter	10.9	14.0	12.6	60.5	96.5	84.3

Cake is removed from the discharge hopper by a system of conveyor belts and is loaded on railroad dump cars for disposal.

9.5.3 Operation and Maintenance

The operating hours of the centrifuges as of September 1986 are as follows:

Unit No.	Operating Hours
1	32,000
2	31,800
3	32,200
4	29,400
5	30,900
6	29,000
7	30,200
8	30,600
9	30,300
10	31,200
11	30,300
12	15,200

In general, the scrolls are repaired after 30,000 hours of operation at a cost of $12,000-$17,000. In 1986, four units had the scrolls repaired. As of 1986, the West-Southwest units have not had the bearings replaced. The conveyor bearings were replaced after 10,000-15,000 hours at a cost of $300.

The 1986 maintenance costs, shown below, reflect an average feed rate of 9.46-11.8 l/s (150-189 gpm) @ 3.51 percent TSS and 88.7 percent recovery.

	Machines Only	Support Equipment	Total
$/hr	4.86	3.21	8.17
$/dry ton	3.81	2.46	6.27

Operating costs, based on $0.053/kWh, sludge feed concentration of 3.51 percent TSS, and 88.7 percent recovery, were as follows:

Flow, gpm	kWh/gpm	$/hr	$/ton DS
180	0.412	3.50	2.80

The 1986 chemical dosage and cost is shown below:

Polymer	lb/dry ton	$/lb	$/ton DS
2540 C	230	0.0528	12.14

The 1986 sludge dewatering costs were estimated to be:

Sludge Dewatering	Cost/Unit	$/ton DS
Polymer @ 230 lb/ton DS	$0.053/lb	12.14
Power @ 52.8 kWh/ton DS (centrifuge) @ 23.0 kWh/ton DS (support eq) @ 75.8 kWh/ton DS (total)	$0.053/kWh	4.02
Water @ 0.22 gph/hr (centrifuge) @ 0.10 gph/hr (support eq) @ 0.32 gph/hr (total)	$0.05/1,000 gal	0.04
Labor @ 0.084 MH/hr (centrifuge) @ 0.092 MH/hr (support eq) @ 0.176 MH/hr (total)	$16.45/MH	2.04
Maintenance		6.27
Total Dewatering Cost		24.51

During May of 1986, the MSDGC tested a pilot screw press (manufactured by Hoilim Iron Works Company, Ltd.) on anaerobically digested sludge at the West-Southwest STP. The test results and screeen press are described in detail in Section 7.6 of this Manual.

Recently MSDGC has conducted dewatering studies on a new type of centrifuge, the Humboldt CentriPress CP2-1. The tests produced the following average results:

Cake -	29.4%
Recovery -	92.7%
Polymer Cost -	$12.15/ton

The complete test data for the Chicago tests are included in Section 7.6 of this manual.

9.6 Case Study, Centrifuges, Denver, CO

The following report was excerpted from a 1986 paper entitled "Digested Sludge Thickening and Dewatering at the Metropolitan Denver Central Plant" (1). In 1981, a study conducted for the Metropolitan Denver Sewage Disposal District No. 1 (MDSDD No. 1) recommended a solids handling system that included centrifugal dewatering and on-site composting of the digested sludge, or alternatively, thickening the waste sludge for agricultural reuse. A solid-bowl centrifuge was installed in 1982 and continuously operated reliably, leading the District to purchase a second, identical unit in 1985. Installation of the centrifuge substantially reduced labor requirements at the plant and caused a reduction in the unit cost of processing the waste sludge.

Wastewater flow to the Central Plant averages 6.8 m^3/s (155 mgd), generating approximately 6.2 Mg (70 dry tons)/d of digested waste solids. The solids are a combination of primary and waste-activated sludges, which are anaerobically digested before disposal. The digested sludge is either thickened for agricultural reuse or dewatered for composting, using two high-capacity centrifuges.

The decision to purchase centrifuges for dewatering was made after on-site testing of two belt filter presses and two centrifuge units during late 1978 and early 1979. From these tests, District personnel concluded that centrifugation had the lowest overall costs. Test results and costs are summarized in Table 9-5 and are compared to the vacuum filter units, which were in operation at that time.

The recommendation for centrifugation was based on the following:

- Centrifuges have lower capital, operation, and maintenance costs.
- Centrifuges had proved higher capacity for peak loads of sludge.
- Centrifuges required only polymer for conditioning, compared to belt presses, which needed both polymer and ferric chloride.
- Centrifuges have higher throughput capacities than belt presses.
- Centrifuges can both thicken and dewater.

The centrifuge pre-purchase specification was directed to both high and low speed units and included a bid evaluation procedure. The bidders were required to include a guaranteed loading (in terms of solids and throughput), power consumption, and polymer dosages. The equipment comparison for dewatering is summarized in Table 9-6.

Based on the data presented in each of the bids, an economic evaluation was completed using the following parameters:

- Electrical energy at current local rates over a period of five years.
- Polymer use at current costs to MDSDD No. 1 in quantity lots over a period of five years.
- Throughput per unit and resultant cake solids.
- Any significant differences in installations and support system costs.
- Any exceptions taken to the design requirements or specifications that might in any way affect the use of this equipment.

On the basis of the evaluation procedure and the equipment costs, a Humboldt-Wedag centrifuge was

Table 9-5. **Comparison of Dewatering Systems - Metropolitan Denver Central Plant[1]**

Item	Belt Press		Centrifuge		Existing Vacuum Filter
	A	B	A	B	
Capital Recovery					
4-yr (1983), $/Mg	7.56	13.61	13.72	12.36	-
10-yr (1989), $/Mg	3.03	5.45	5.49	4.95	-
Chemicals, $/Mg	48.03	44.25	24.16	20.13	54
Operations Labor, $/Mg	6.47	6.47	4.31	4.31	6.47
Power, $/Mg	0.22	0.33	2.28	3.88	1.58
Water, $/Mg	2.07	1.93	-	-	0.36
Total[2], $/Mg	67.38	72.04	49.96	45.63	62.41
Cake Solids, percent	17.5	17.0	14.0[3]	12.03	9.5
Solids Recovery, percent	91	91	90-95	94-95	75-80

[1] Based on processing 86.9 Mg/d.
[2] Based on 4-yr capital recovery value
[3] Based on these results, plant personnel predicted a cake solids of 16 percent.

Table 9-6. **Centrifuge Bids - Metropolitan Denver Central Plant**

	Bid Specification		Humboldt-Wedag		Unit A		Unit B	
			$530,693		$366,000		$630,500	
Design Requirements	Dewatering	Thickening	Dewatering	Thickening	Dewatering	Thickening	Dewatering	Thickening
Volatile Matter, %	61	61	-	-	-	-	-	-
Guaranteed Loading								
dry tons/d*	-	-	96	144	57.72	115.4	60	120
dry lb/hr*	-	-	8,000	12,000	4,810	9,620	5,000	10,000
gpm*	-	-	500	750	300	600	300	600
Min. Feed Rate, gpm	300	600	-	-	-	-	-	-
Feed Conc., %TS	2.5-3.2	2.5-3.2	-	-	-	-	-	-
Min. TS Recovery, %	90	90	-	-	-	-	-	-
Min. Cake Solids, %	16	6	-	-	-	-	-	-
Max. Polymer, lb/ton*	-	-	10.0	2.5	13	4-5	15	8
Max. HP Draw, hp/gpm*	-	-	0.27	-	120	171	220	220

* Values supplied by bidders.

selected. The cost comparison is summarized in Table 9-7 and shows that the evaluated cost of the Humboldt machine is 19 percent less than Unit A and 49 percent less than Unit B for solids dewatering.

The centrifuge installed at the Central Plant is a Humboldt-Wedag S6-1, and it is believed to be the largest centrifuge in daily use in a municipal facility in the U.S. This concurrent unit has a bowl dimension of 1.4 m x 4.3 m (56 in x 168 in) long. The bowl is driven by a 149-kW (200-hp) main drive motor and the scroll by a 56-kW (75-hp) hydraulic backdrive system. The backdrive is controlled by a microprocessor (see Figure 9-8). During automatic operation, a user-programmable control curve is keyed into the microprocessor. The control curve defines the relationship between the scroll conveyor torque (measured as hydraulic pressure) and the differential speed. Figure 9-9 shows a control curve in which the torque is measured as hydraulic pressure at the inlet to the backdrive hydraulic motor. The operating point on the curve is a function of the slurry feed rate and the feed solids concentration. The control point is set up with a "warning" torque set point, which, when reached, will cause the backdrive to increase the differential speed to its maximum. This increase clears solids from the machine and thereby avoids an automatic shutdown.

The centrifuges installed at the Central Plant have been modified, based on research completed by Humboldt-Wedag. The angle and spacing of the scroll conveyors have been changed to optimize detention time in the centrifuge. Another change was to raise the maximum torque delivered to the backdrive hydraulic motor. The higher torque results in larger throughput and drier cakes.

The centrifuge was started in May 1982 and test results are summarized in Table 9-8. The Denver

Table 9-7. Centrifuge Evaluation for Dewatering and Thickening - Metropolitan Denver Central Plant

Item	Five-Year Cost (1982$) Dewatering	Thickening
Unit A		
Capital	366,000	366,000
Polymer	2,807,142	1,926,653
Energy	178,809	254,803
Total	3,352,951	2,547,456
Unit B		
Capital	630,500	630,500
Polymer	3,239,010	3,425,160
Energy	327,817	476,824
Total	4,197,327	4,532,484
Humboldt		
Capital	530,693	530,693
Polymer	2,159,340	1,070,363
Energy	120,696	241,392
Total	2,810,729	1,842,448

Assumptions: Costs straight-lined for 5 years
Power - $0.04536/kWh
Polymer - $0.93/kg ($2.04/lb)
Thickening - 136 m^3/hr, 105 Mg/d (600 gpm, 115 dry tons/d)
Dewatering - 78m^3/hr, 53 Mg/d (300 gpm, 58 dry tons/d)

sludge is difficult to dewater; part of the problem is believed to be the preponderance of waste activated sludge in the final product. Ferric chloride is used to adjust the pH of the sludge and avoid scaling (struvite deposition) of piping and equipment.

The centrifuge has performed reliably since its installation and has simplified management and handling of waste sludges at the Central Plant. The success of the centrifuge installation prompted the District to purchase a second, identical machine as a backup.

Higher loading rates became achievable after larger feed pumps were installed. The results of the cake production tests then are presented in Figures 9-10 and 9-11. This also was after old sludge was cleaned out and equilibrium was established. This data shows a substantial improvement over the results of acceptance tests and establishes a broader range of flow rates than were originally maintained under the acceptance testing. The operational performance results in Figure 9-10 show that the centrifuge is capable of achieving greater than 19 percent total solids (TS) with the range of 31-57 l/s (500-900 gpm) flow rate. The cake dryness can be controlled within the narrow range of 19.0-19.4 percent TS, while varying the feed rate through the use of the microprocessor. At differentials of 1 to 9 rpm for the corresponding flow rates of 31-57 l/s (500-900 gpm), the cake dryness and percent capture are 19 and 90 percent, respectively. The polymer feed rate varied from 11 kg/Mg (22 lb/ton) to 16 kg/Mg (32 lb/ton) of sludge, as shown in Figure 9-11. The variation in polymer feed rate was attributed to the inverse relationship with the feed solids concentration (Figure 9-11). The horsepower used per unit of flow decreased with increasing flow (Figure 9-12).

The main feature of this installation is the positive impact on the District's annual operating costs. Table 9-9 compares the 1986 centrifuge operation and maintenance costs (O&M) costs to the last year (1981) of operating the vacuum filters. Table 9-9 shows that, on a unit cost basis, the centrifuge cost is 61 percent of the vacuum filter cost. The cost difference is attributed to the higher centrifuge throughput. Because the centrifuge was able to process all of the sludge (more than twice the quantity processed by the vacuum filters), chemical and electrical costs are higher but the cost per ton processed is lower. The personnel costs reflect one full-time operator per shift (total of three) and a maintenance staff consisting of one mechanic, two utility repairmen, and one electrician. This staff services not only the centrifuge but all support equipment such as conveyors, process controllers, and chemical/sludge feed equipment. Materials costs reflect spare parts used for routine preventive and corrective maintenance. Chemical requirements for polymer are 218 Mg (240 tons)/yr and for ferric chloride are 481 Mg (530 tons)/yr. Outside services are primarily for rebuilding the centrifuge bowl and scroll ($45,000).

The digested sludge is dewatered (19 percent TS) for composting on-site, or, alternatively, thickened sludge (8 percent TS) is transported to agricultural lands where it is injected. The ultimate disposal costs for 1981 and 1983 are summarized in Table 9-10. The centrifuge installation has resulted in a 1986 annual savings of $627,259.

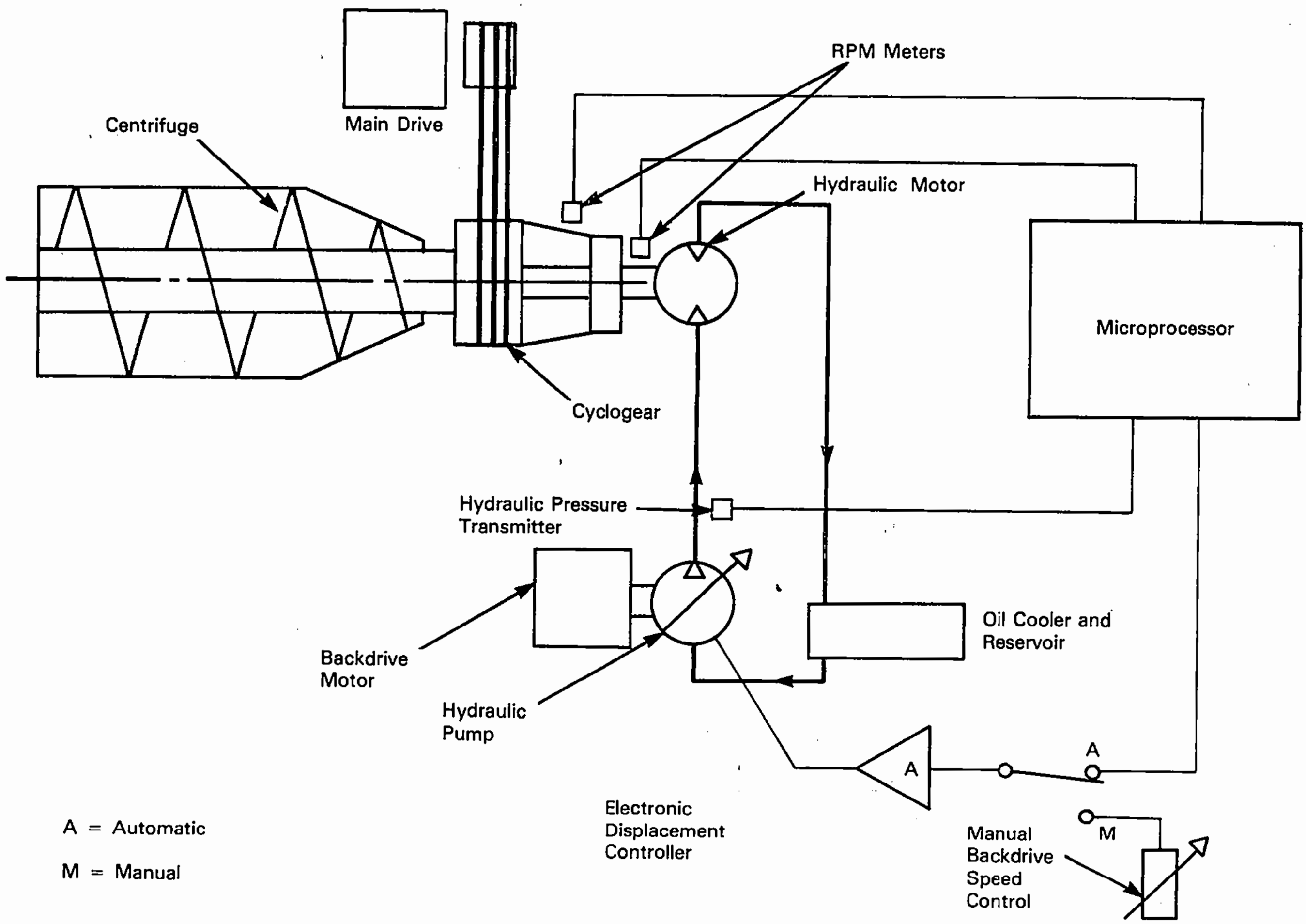
The use of a centrifuge has shown benefits for larger treatment plants, when compared to continuous belt filter presses. These larger units require less space, less operator attention, and have a lower initial cost. The Central Plant would have required at least six 2-m belt presses to achieve the same capacity as one centrifuge. Based on the earlier on-site testing, the cake solids would not have exceeded that produced by this centrifuge.

9.7 Case Study, Centrifuges, Ontario, Canada

9.7.1 Sludge Characteristics and Processing

The Duffin Creek Water Pollution Control Plant (municipalities of York and Durham) is a 1.8-m^3/s (40-mgd) conventional activated sludge facility. It is designed for the removal of phosphorus in accordance with the Canadian requirements for the

Figure 9-8. Automatic backdrive system.



protection of the Great Lakes. Ferrous sulphate is added to the aeration system for the removal of phosphorus. Design features of the plant include:

Primary clarifiers:
4 @ 24 m x 23 m x 3.7 m SWD
(80 ft x 77 ft x 12 ft)

Aeration basins:
16 @ 23 m x 23 m x 5.8 m SWD
(75 ft x 75 ft x 19 ft)

Secondary clarifiers:
8 @ 41.2 m dia. x 3.7 m
(135 ft dia. x 12 ft)

The 1986 operational characteristics of the plant were as follows:

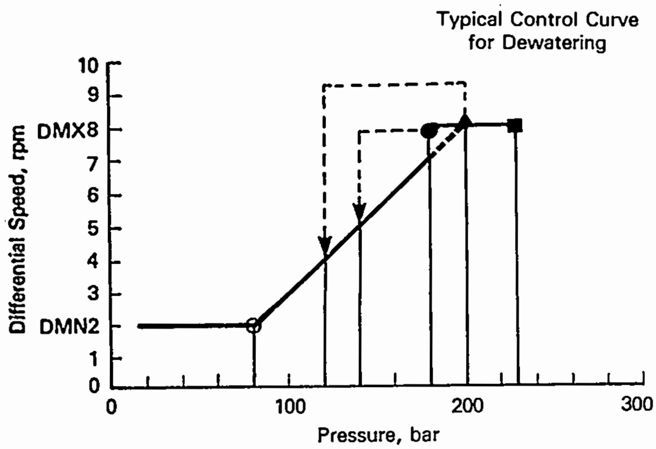
	Influent	Secondary Effluent
Flow, mgd	35.6	35.6
BOD_5, mg/l	147	20.8
TSS, mg/l	266	17.8
TP, mg/l	6.0	0.9

The plant was designed on the basis of an F/M of 0.3 and a Mean Cell Retention Time (MCRT) of 8 days. The actual operating conditions for 1986 were as follows: 0.2 F/M and 15 days MCRT. A schematic of the Duffin Creek Pollution Control Center is shown in Figure 9-13.

At Duffin Creek, the excess waste activated sludge and the primary sludge are digested in conventional digesters with the designed retention time of 30 days. The blend sent to the digesters is approximately 70 percent WAS and 30 percent PS. The material, at present, is digested for 20 days and then dewatered at a concentration typically between 4 and 6 percent solids. The daily quantity varied between 15 Mg/day and approximately 19 Mg/d in 1986.

The waste activated sludge system is presently producing 45,450 kg WAS/d and the primary clarifier is capturing 38,290 kg suspended solids/d. The waste, RPS + WAS, is sent to digestion after blending. The SVI averaged approximately 60 ml/g in 1986, and varied between a low of 35 and a maximum value of 78 ml/g.

Figure 9-9. Typical control curves for automatic backdrive.

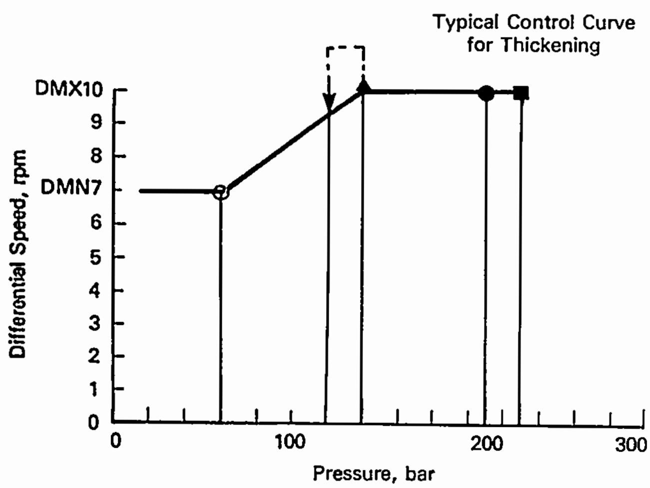


bar x 100 = kPa.

- O = First Control Point
- ● = Second Control Point, Reduce to 0.8 x Control Point
- ▲ = Maximum, Reduce to 120 Bar
- ■ = Automatic Shutdown

DMX = Maximum Differential Speed

DMN = Minimum Differential Speed



bar x 100 = kPa.

- O = First Control Point
- ▲ = Maximum, Reduce to 120 Bar
- ● = Second Control Point
- ■ = Automatic Shutdown

DMX = Maximum Differential Speed

DMN = Minimum Differential Speed

Table 9-8. Comparison of Bid Specifications and Actual Performance for the Dewatering Mode

Item	Bid Specification	Actual Performance	
		Average	Range
Feed Rate, gpm	500.0	424.0	200-600
Feed Conc. %TS	2.5-3.2	2.6	2.3-3.5
Solids Recovery, %	90.0	89.0	82-95
Polymer Dose, lb/ton	10.0	14.2	10.4-19.8
Ferric Dose, lb/ton	-	47.0	25-70
HP Draw, hp/gpm	0.27	0.37*	

* 0.32 hp/gpm @ 800 gpm

Figure 9-10. Centrifuge dewatering.

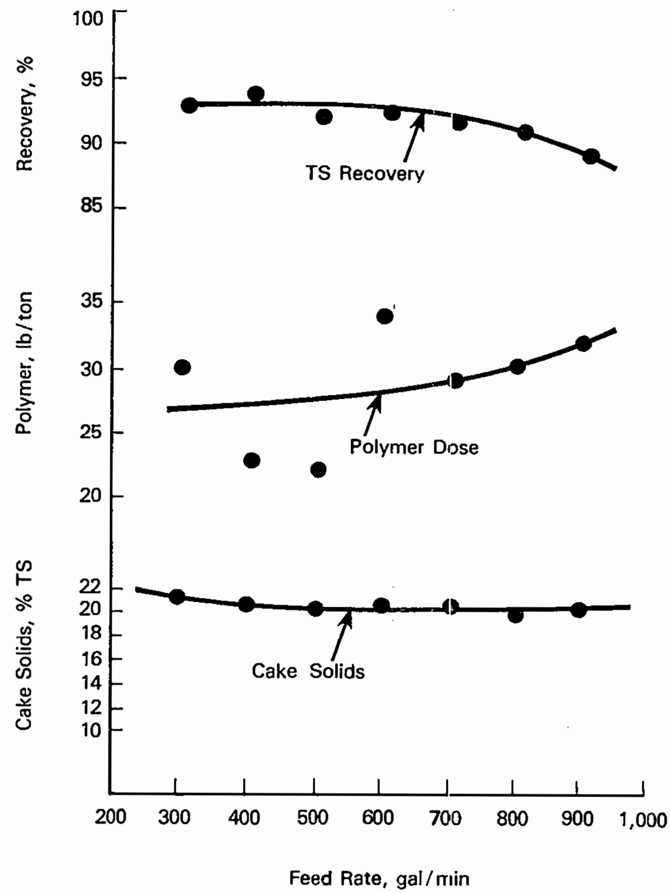


Figure 9-11. Polymer requirements.

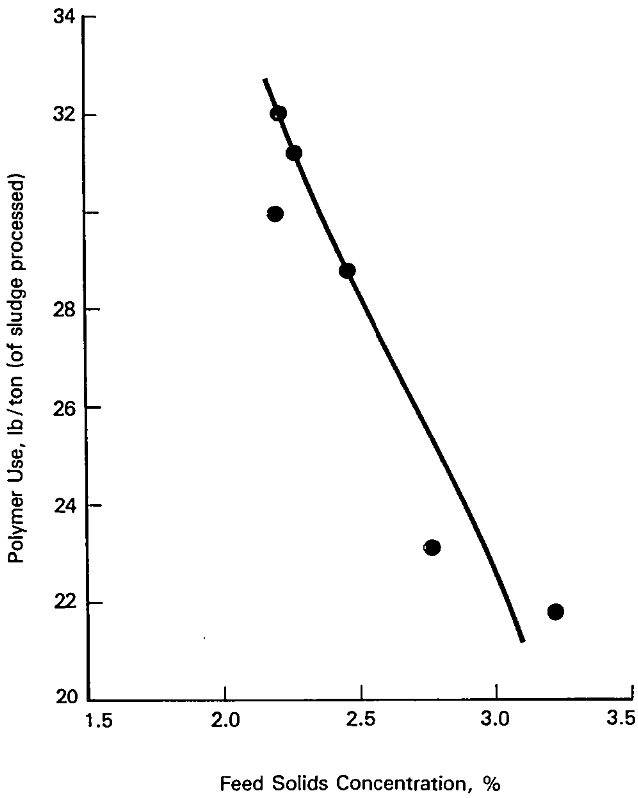
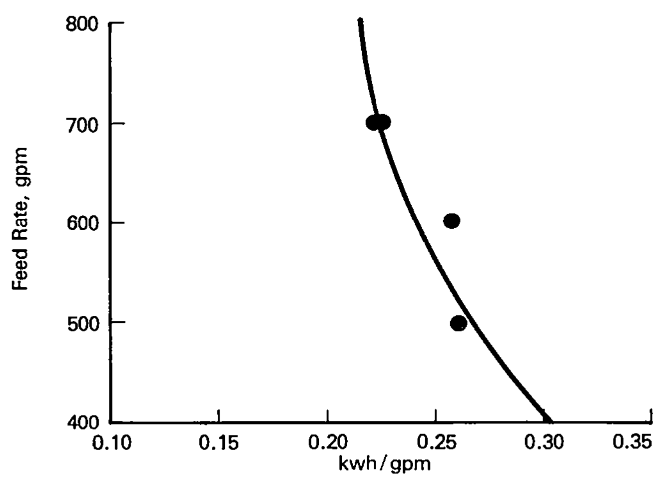


Figure 9-12. Centrifuge power curve (dewatering mode).



9.7.2 *Filter Presses*

Four membrane filter press systems are used for dewatering sludges. Each filter is currently equipped with 66-1,200 mm x 1,200 mm plates. The system, however, has been expanded to 83 plates. Sludge is fed to the units with positive displacement pumps manufactured by Thomas Willett & Co., Ltd. A measured amount of polymer is injected into the

Table 9-9. **Dewatering Equipment Comparison - Metropolitan Denver Central Plant**

Activity	Vacuum Filter (1981$)	Centrifuge (1986 $)
	$/day	$/day
Personnel	766.50	1,173.04
Materials	166.61	390.80
Chemicals	1,199.76	2,659.36
Electricity	154.57	299.45
Outside Services	19.31	130.23
Total	2,306.75	4,652.88
Inflation (5%/yr)	2,944.75	
Sludge, tons/d	29	75
Unit Cost	101.52	62.04

Table 9-10. **Ultimate Disposal Costs - Metropolitan Denver Central Plant**

Activity	Vacuum Filter (1981$)	Centrifuge (1986 $)
	$/day	$/day
Personnel	4,566.33	4,944.56
Materials	2,167.02	1,297.64
Chemicals	1,199.76	2,659.36
Electricity	174.36	363.60
Outside Services	600.44	130.23
Total	8,707.91	9,395.39
Inflation (5%/yr)	11,113.91	
Sludge, tons/d	73	75
Unit Cost	152.25	125.27

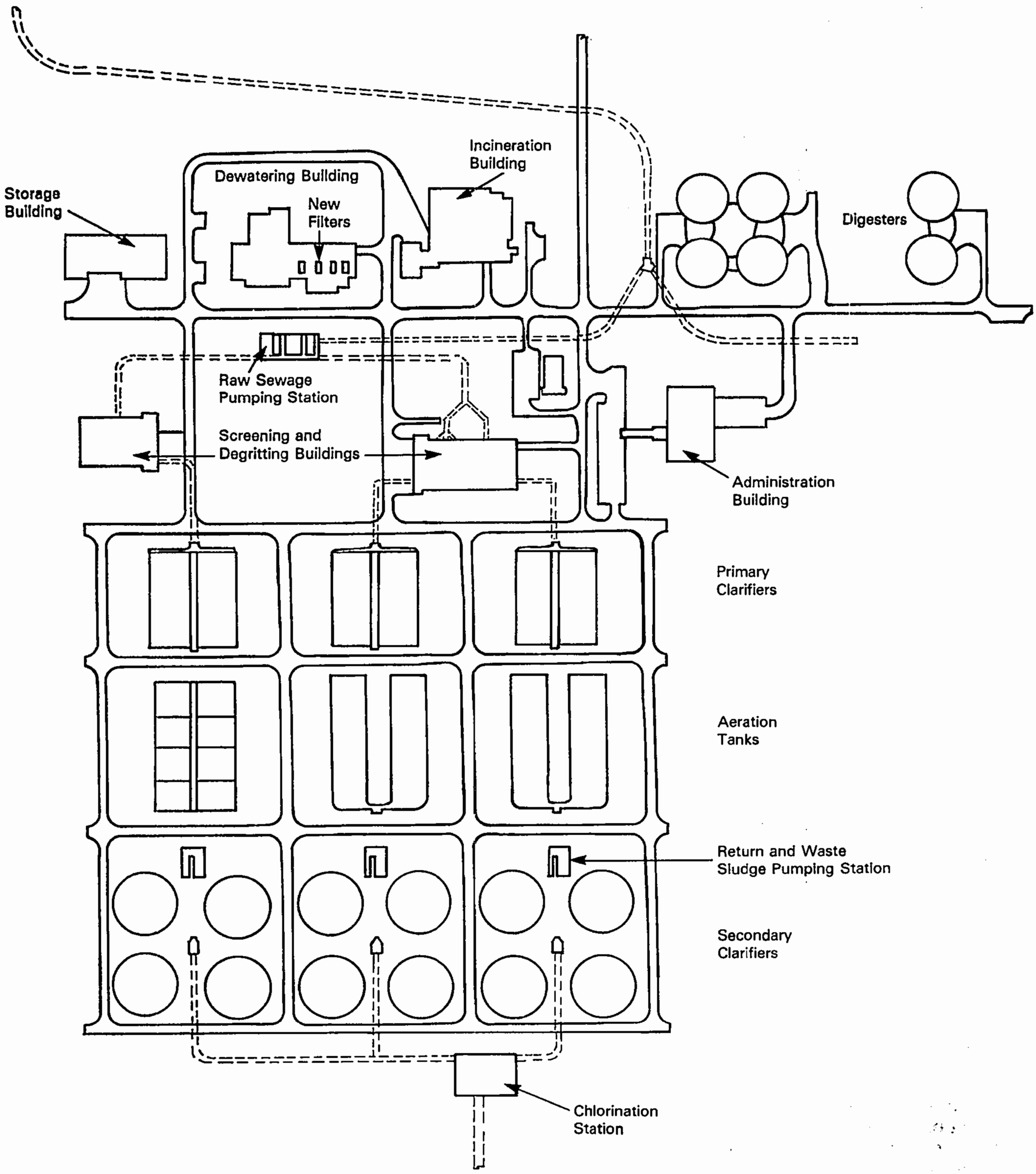
Willett pump on suction stroke and returns to the storage tank on the discharge stroke. Polymer is taken from the storage tank by a Moyno pump and sent to the Willett pump. Electric solenoids determine the end destination, that is, in pump or in tank. This system is very reliable for delivering a constant volume. It is called a "Polymeter System" and is supplied by Allied Colloids.

The sludge is conditioned with polymer, but without precoat. A polymer dissolving system is used at Duffin Creek. The humidity is sufficiently low in the Toronto area so the hydroscopic properties of the polymer have not adversely affected the dewatering operation.

Augers and screws are used to convey all sludges at the Duffin Creek facility. The augers, manufactured by Asdor, Ltd., are installed in 4.6 m sections and are coupled with flexible connectors and grease fittings.

The filter presses at Duffin Creek have been in operation since 1985. The filter presses replaced belt presses, installed in 1981, which produced only 18-21 percent TS. The plant encountered the normal

Figure 9-13. Schematic of Duffin Creek Pollution Control Plant, Ontario, Canada.



start-up difficulties; however, they were compounded by the discharge of unusually large quantities of difficult-to-filter waste activated sludge. Although digested waste activated sludge is currently fed to the presses, in the beginning the presses successfully dewatered raw sludge to 28-30 percent cake solids. Actual daily data for the months of September and December 1986 is compiled in Tables 9-11 and 9-12, respectively. The data indicates that the solids content of the cake exceeded the 30 percent requirement during most of the period. In 1986, this trend continued during September through December as shown in the monthly summaries in Table 9-13. Table 9-14 presents 1986 operating data for the filter press operation.

Table 9-11. Duffin Creek WPCP Dewatering System (September 1986 Operating Data)

Date	Cycles/day	Cake Dry Solids	Total Dry Solids	Dry Solids/Cycle
September		percent	Mg/d	Mg
1	36	31.2	35.06	0.97
2	36	32.5	36.74	1.02
3	33	33.1	33.95	1.03
4	34	31.5	33.28	0.98
5	40	32.4	39.36	0.98
6	28	30.8	26.33	0.94
7	18	29.8	16.23	0.90
8	22	32.3	21.50	0.98
9	40	34.4	43.28	1.08
10	31	31.6	30.69	0.99
11	37	32.9	38.09	1.03
12	35	33.1	35.54	1.02
13	21	34.8	21.90	1.04
14	23	34.1	24.55	1.07
15	37	34.0	37.19	1.01
16	36	31.8	34.16	0.95
17	12	32.0	11.94	1.00
18	38	33.0	38.36	1.01
19	40	33.8	41.64	1.04
20	29	33.3	29.56	1.02
21	19	34.0	20.35	1.07
22	8	32.0	7.76	0.97
23	32	32.6	31.69	0.99
24	34	34.0	36.16	1.06
25	39	35.1	42.83	1.10
26	31	35.6	33.12	1.07
29	20	33.0	20.86	1.04
30	21	34.3	23.32	1.11
Average	30	33.0	30.19	1.02

The system, as indicated, uses only polymer. Precoat is not employed. In general, the operators indicate the cake release is fair to excellent. However, they have to observe the cake discharge to ensure it is complete.

The operating cycle time is typically 90-95 minutes, which includes 45 minutes of fill and 50 minutes of squeeze time. The average core blow time is 25 seconds. The core blow, however, is incomplete, and consequently core solids restrict the cake discharge. The poor core discharge is partially linked to the extremely high solids content of the core. In addition, the pipe design at the core blow line restricts the solids. To improve cake discharge, the core blow receiving pipes must be larger than the core diameter and must not make any abrupt changes in direction.

Table 9-12. Duffin Creek WPCP Dewatering System (December 1986 Operating Data)

Date	Cycles/day	Cake Dry Solids	Total Dry Solids	Dry Solids/Cycle
December		percent	Mg/d	Mg
1	30	34.6	32.49	1.08
2	32	35.3	34.70	1.08
3	37	34.0	39.26	1.06
4	36	33.5	37.99	1.06
5	36	34.0	38.56	1.07
6	22	34.0	23.61	1.07
7	21	34.0	21.97	1.05
8	34	34.6	36.62	1.08
9	33	36.5	37.42	1.13
10	34	38.8	42.39	1.25
11	35	37.5	41.34	1.18
12	26	36.0	29.07	1.12
13	28	37.3	32.15	1.15
14	21	35.8	23.46	1.12
15	35	37.1	41.53	1.19
16	35	33.4	36.24	1.04
17	35	34.3	37.39	1.07
18	33	34.0	35.31	1.07
19	33	33.9	34.94	1.06
20	16	34.0	17.01	1.06
21	16	37.3	20.32	1.20
22	32	30.3	29.88	0.93
23	33	31.2	31.61	0.96
24	7	30.9	6.75	0.96
25	8	34.8	9.11	1.14
26	25	34.3	26.80	1.07
27	28	36.0	31.81	1.14
28	17	31.3	16.64	0.98
29	35	32.5	35.12	1.00
30	30	37.3	35.41	1.18
31	27	31.9	26.52	0.98
Average	28	34.5	30.43	1.08

The typical polymer dose is 5 to 6 kg/Mg (10-12 lb/ton) of 100 percent active material, with a low dose of 3 kg/Mg (6 lb/ton) and a high of 8 kg/Mg (16 lb/ton). Allied Colloids' Percol 757 is used at a cost of $2.80/kg dry powder. In recent months, the ability of the diaphragm plate presses to achieve 33-36 percent TS on digested sludge has resulted in an autogenous operation of the fluid bed reactors. At 18-21 percent TS, the furnace capacity was reduced 40 percent and fuel usage was over 400 l/Mg dry solids. The furnace was designed to receive a minimum of 30% TS from the belt presses.

Table 9-13. **Duffin Creek WPCP Sludge Loading for 1986**

	Raw Sludge			Digested Sludge			Presses		
	% solids	% volatile	volume, m^3	% solids	% volatile	volume, m^3	% solids	% volatile	volume, m3
January	4.25	56.33	32,045.2	4.61	47.85	10,187.9	26.47	43.84	2,135.5
February	4.63	58.78	23,462.1	3.63	49.49	9,120.4	26.47	44.12	1,898.9
March	4.77	57.18	24,971.2	4.11	46.84	14,829.9	26.00	49.95	2,986.8
April	5.15	61.80	24,443.0	4.60	47.28	14,706.9	27.64	45.83	2,755.1
May	4.90	63.44	37,544.3	4.10	55.75	18,627.2	26.17	47.10	3,025.3
June	5.40	60.97	24,608.0	5.38	49.00	13,611.9	26.20	45.83	2,677.1
July	5.23	55.80	34,322.0	5.86	45.60	12,970.7	27.84	44.21	2,883.3
August	7.09	50.69	33,120.0	6.73	42.50	14,868.3	29.40	40.06	2,851.0
September	7.40	42.60	27,530.0	8.25	36.18	21,143.0	32.90	35.70	2,659.6
October	6.04	51.43	38,858.0	10.00	34.94	6,982.0	32.11	32.69	2,129.0
November	5.97	52.54	22,448.0	8.14	36.15	8,542.0	35.06	32.54	1,968.2
December	5.25	53.00	25,096.0	7.04	35.77	14,223.0	34.59	34.94	2,725.5
Average	5.51	55.38		6.04	43.95		29.24	41.40	
Total			348,447.8			159,813.3			30,695.5

Table 9-14. **Duffin Creek WPCP Sludge Dewatering for 1986**

	Volume	% Solids	CST	Cake Solids
		Mg	sec	%
January	571.66	4.2	465.2	26.7
February	503.62	4.0	417.7	26.6
March	787.60	4.1	384.6	26.4
April	759.13	4.2	449.4	27.6
May	809.92	3.8	522.6	26.6
June	682.77	3.9	575.9	26.1
July	800.62	4.3	554.0	27.8
August	835.64	5.3	506.8	28.1
September	878.47	7.3	587.4	32.9
October	678.09	7.4	511.9	31.9
November	683.91	8.1	537.3	34.9
December	943.42	7.0	481.4	34.5
Average	744.57	5.3	499.52	29.2
Total	8,934.85			

9.7.3 Operation and Maintenance

After 10,880 operating hours, the cloths were changed for the first time. Prior to the cloth change, filtration rates and cake solids had remained at a satisfactory level.

The gritty character of the digested sludge has created operating and maintenance problems at Duffin Creek. Fine grit passes through the filter medium and lodges in the plates; these plugged points hinder cake release. As a result, the plant removed the cloths after five months and cleaned the entrapped fines. After another four months, the plant replaced the cloths with new ones. Recently, they have changed the type of cloth from RILSAN to polypropylene and now plan to change the cloths every six months.

In general, the operators like the filter press operation at Duffin Creek and generally feel it is a cleaner operation than the original belt press installation. One exception would be when blowout occurs and the cover screens are not in place. The system, however, is now equipped with "fail-safe" curtains (presently provided by ISB in Montreal), which provide a warning signal and automatically stop the press cycle if the safety screens are not in place.

The initial operation used plant air for both instruments and the compression cycle. As a result, the instrument part of the operation was starved for air during the compression periods. To correct the problem, a second compressor was installed specifically for the instruments.

The operators feel the polymer system is fundamentally a good one. However, as the proportion of primary to secondary sludge varies (even after digestion), the polymer dosage will also change, thus affecting the cake release. An ideal situation would be to supply consistent quality of sludge to the unit.

9.8 Reference

1. Williams, R.B. and J.K. Nelson. *Digested Sludge Thickening and Dewatering at the Metropolitan Denver Central Plant*. Presented at the 59th Annual Conference of the Water Pollution Control Federation, Los Angeles, CA, October 5-9, 1986.

Appendix A
Design Examples/Cost Analyses

A.1 Introduction

This appendix presents design examples for several dewatering processes. The examples are prepared for two different sizes of treatment plants: 0.088 m^3/s (2 mgd) and 0.88 m^3/s (20 mgd). Information from the design chapters is used as the basis for the calculations. Cost analyses are also presented to show how to prepare cost estimates by using the EPA publication *Handbook: Estimating Sludge Management Costs* (1).

Note: Construction cost estimates presented in this Appendix are total base capital costs obtained from the EPA Handbook (1). Total base capital costs (TBCC) for sludge dewatering processes presented in this Appendix include structural, mechanical, equipment, electrical, and instrumentation costs. They do not include costs for engineering design, construction supervision, legal costs, administration, interest during construction, and contingencies. In order to estimate the total project construction cost, these non-construction costs must be estimated and added to the process TBCC costs derived from the cost curves in the EPA Handbook. In addition, the costs presented in this Appendix are based on last quarter 1984 costs, and must be adjusted for inflation for use in later years. An example of how to estimate the total project construction cost given the TBCC is shown for upgraded sand drying beds under Section A.3.2 for the 0.088-m^3/s (2-mgd) plant. More detail on these cost update procedures can be found in Section 2.6 of the EPA Handbook (1).

A.2 Determine Sludge Quantities

Assume Activated Sludge Plant with Primary Clarifiers

Sludge Type:

Primary Sludge (P) + Waste Activated Sludge (WAS)

Sludge Quantities:

P - 150 kg solids/Ml treated (1,250 lb/Mgal)
WAS - 90 kg solids/Ml treated (750 lb/Mgal)

Sludge Solids Concentrations:

P - 5.0% from primary clarifier
WAS - 0.5% from secondary clarifier
Thickened WAS - 4.0% from dissolved air flotation thickener

Determine Sludge Volumes Before Anaerobic Digestion

Assume Sludge Solids Specific Gravity

= 1.4 for primary sludge
= 1.25 for WAS

Sludge Specific Gravity [Equation 2-3 from (1)]:

$$SSG = \frac{1}{\dfrac{SS}{(100)(SPG)} + \dfrac{(100 - SS)}{100}}$$

where,

SSG = sludge specific gravity (dimensionless).
SS = sludge suspended solids concentration, weight percent.
SPG = sludge solids specific gravity (dimensionless).

Primary Sludge SSG
= 1/[[(5)/(100)(1.4)] + [(100 - 5)/100]] = 1.01

Thickened WAS SSG
= 1/[[(4)/(100)(1.25)] + [(100 - 4)/100]] = 1.01

Therefore, combined P + WAS has SSG = 1.01.

Sludge Volume:

$$SV = \frac{(DSS)(100)}{(SS)(1.0\ kg/l)(SSG)}$$

where,

SV = sludge volume, l/Ml treated
DSS = dry sludge solids produced, kg/Ml
SS = sludge suspended solids concentration, percent
SSG = sludge specific gravity (dimensionless)
1.0 = density of water, kg/l

SV = [(150 kg/Ml treated)/(0.05)(1.0)(1.01)]
 + [(90 kg/Ml treated)/(0.04)(1.0)(1.01)]
 = 5,200 l/Ml treated (5,200 gal/Mgal)

Solids Content before Digestion
 = (100)(150 + 90)/[(5,200)(1.0)(1.01)] percent
 = 4.6 percent solids

Assume anaerobic digestion destroys 50% of volatile solids. Assume 70% of sludge solids are volatile.

Mass of solids before digestion:
150 + 90 = 240 kg/Ml

Solids Destroyed = (240 kg/Ml)(0.70)(0.50)
 = 84 kg/Ml (700 lb/Mgal)

Solids Remaining = 240 kg/Ml - 84 kg/Ml
 = 156 kg/Ml (1,300 lb/Mgal)

Solids Content after Digestion
 = (100)(156 kg/Ml)/(5,200)(1.0)(1.01) percent
 = 3.0 percent solids

A.3 Upgraded Sand Drying Beds

A.3.1 Design Examples

When designing sand drying beds for a wastewater treatment plant, the required size or bed area depends on the geographic location of the plant. In an area with a high evaporation rate, the bed area can be smaller than in an area with a low evaporation rate.

From Chapter 6 of this manual, typical loading criteria for anaerobically digested primary sludge plus waste activated sludge are 60 to 100 kg/m^2/yr (12 to 20 lb/sq ft/yr). For polymer conditioned sludges these criteria can reasonably be increased by 50 to 100 percent to 120 to 200 kg/m^2/yr (25 to 41 lb/sq ft/yr). For this example, use 140 kg/m^2/yr (29 lb/sq ft/yr). This loading rate would be appropriate for southwestern and southern regions of the United States. For northern regions of the United States, a loading rate of 80 to 100 kg/m^2/yr (17 to 21 lb/sq ft/yr) is more appropriate.

Design Example for a 0.088-m^3/s (2-mgd) Plant

Sludge Quantity

Digested Sludge Solids
= (0.088 m^3/s)(156 kg/Ml)(86,400 s/d)/(1,000 m^3/Ml)
= 1,190 kg solids/d (2,600 lb/d)

Digested Sludge Volume
= (1,190 kg/d)/(0.03)(1.0 kg/l)(1.01)
= 39,300 l/d (10,400 gpd)

Determine Required Sand Bed Area (based on solids loading from above)

Bed Area = (1,190 kg/d)(365 d/yr)/140 kg/m^2/yr
 = 3,100 m^2 (33,000 sq ft)

Check Number of Sludge Applications per Year [assume normal application depth of 23 cm (9 in)]

Total Volume Available per Application
 = (3,100 m^2)(23 cm)(1,000 l/m^3)/100 cm/m
 = 713,000 l (188,000 gal)

Number of Applications Per Year
 = (39,300 l/d)(365 d/yr)/(713,000 l/application)
 = 20 applications/yr

If the plant is located in the northern part of the United States, 20 applications per year is too high for design purposes. In the southern and southwestern United States, this loading rate is reasonable.

Determine Number of Beds

If beds are 7.6 m wide by 30.5 m long (25 ft by 100 ft), the area of one bed is 232 m^2 (2,500 sq ft).

Approximate No. Beds = 3,100 m^2 /(7.6m)(30.5m)
 = 13.4 beds, use 14 beds

Fix Width @ 7.6 m (25 ft)

Length = 3,100 m^2/(14)(7.6 m) = 29.1 m (95.6 ft)

Design Example for a 0.88-m^3/s (20-mgd) Plant

The design approach is the same as in the previous example with the exception that larger sand beds, such as 30 m x 60 m (100 ft x 200 ft), would be used. The overall bed area will be determined.

Sludge Quantity

Digested Sludge Solids
= (0.88 m^3/s)(156 kg/Ml)(86,400 s/d)/(1,000 m^3/Ml)
= 11,900 kg/d (26,000 lb/d)

Digested Sludge Volume
= (11,900 kg/d)/(0.03)(1.0 kg/l)(1.01)
= 393,000 l/d (104,000 gpd)

Determine Sand Bed Area Required

Bed Area = (11,900 kg/d)(365 d/yr)/140 kg/m^2/yr
 = 31,000 m^2 (330,000 sq ft)

A.3.2 Cost Analyses

Cost Analysis for a 0.088-m^3/s (2-mgd) Plant

Determine Capital Cost

Refer to Figure 5-13 on page 85 of the EPA *Handbook: Estimating Sludge Management Costs* (1). Note: English units only are used in the EPA Handbook.

Need Sludge Volume per Year (from Design Example):

10,400 gpd x 365 d/yr = 3.80 x 10^6 gal/yr

Figure 5-13 is based on sludge loading rates of 15 lb/sq ft/yr at 2 percent solids and 22 lb/sq ft/yr at 4 percent solids, or about 18.5 lb/sq ft/yr at 3 percent solids. The design example size was based on 29 lb/sq ft/yr. Using a curve based on a low loading rate to size the sand bed would yield a bed (and cost) which is too large.

Adjust sludge volume by ratio of 18.5/29 = 0.64:

0.64 x 3.8 Mgal/yr = 2.4 Mgal/yr

From Figure 5-13 @ 3% solids:

Capital Cost = $140,000

Note: This includes a land cost valued at $3,120 per acre. If a significantly different cost for land is used, the capital cost needs to be adjusted, as shown in Section 5.8.1 of the EPA Handbook (1).

Note: The drying beds in the EPA Handbook (1) have no plastic or asphalt lining beneath the beds for groundwater protection and no concrete tracks for easy equipment access. If the user elects to include a PVC liner, add about $2.50/sq ft. If the user elects to include concrete tracks, add about $2.00/sq ft.

<u>Example - Convert capital cost to total project construction cost</u>

Update cost to May 1987.

Engineering News Record (ENR) construction cost index in May 1987 is 4367. ENR index for last quarter 1984 is 4171:

4367/4171 = 1.047

Updated capital cost = $140,000 x 1.047
= $146,600

Add non-construction costs to updated cost:

Engineering Design @ 10% of $146,600
= $14,700

Construction Supervision @ 5% of $146,600
= $7,300

Legal and Administrative Costs @ 20% of $146,600
= $29,300

Contingencies @ 15% of $146,600
= $22,000

Subtotal Cost = $219,900

Interest During Construction (assume 10% Interest and 1-year Construction Period):

0.10 x 1 yr x 1/2 x $219,900 = $11,000

Total Project Construction Cost = $230,900

More detail on these cost update procedures can be found in Section 2.6 of the EPA Handbook (1).

Determine O&M Cost
Use Figure 5-14 of the EPA Handbook. Use actual sludge volume to determine O&M cost.

From Figure 5-14 @ 3.8 Mgal/yr and 3% Solids:

Base Annual O&M Cost = $16,000/yr

Use Figure 5-16 of the EPA Handbook to determine O&M cost components. From Figure 5-16:

Labor: 750 hr/yr x $13.50/hr =	$10,100/yr
Diesel Fuel: 2,500 gal/yr x $1.35/gal =	$3,400/yr
Materials:	$2,000/yr
Total =	$15,500/yr

$15,500/yr is close enough to total O&M of $16,000/yr -- within accuracy of curves.

Determine Costs for Polymer Feed System
A typical polymer cost in $/ton dry solids is $11/ton, from the case studies on air drying systems in Chapter 8. From O&M Cost Figure 6-22 from the EPA Handbook (1), the assumed cost of polymer is $2.80/lb.

Based upon this dry polymer cost, the approximate dry polymer dosage is:

($11/ton)($2.80/lb) = 3.9 lb/ton dry solids

Therefore, use a polymer dosage of 4 lb/ton when determining the capital costs.

Figures 6-19 and 6-20 from the EPA Handbook (1) are used to estimate the capital costs for polymer feed systems treating a sludge with 2 percent and 4 percent solids, respectively. Since the design example has 3 percent sludge solids, an average of the capital costs from each figure will give a useable cost.

Use Figure 6-19, based on 3.8 Mgal/yr and 2% solids:

Capital Cost = $31,000

Use Figure 6-20, based on 3.8 Mgal/yr and 4% solids:

Capital Cost = $33,000

For 3% solids sludge, use an average of $32,000 for the capital cost.

For the O&M cost, similarly:

Figure 6-22 gives an O&M cost of $14,000/yr.
Figure 6-23 gives an O&M cost of $20,000/yr.

For 3% solids sludge, use an average of $17,000/yr for the O&M cost.

Cost Analysis for a 0.88-m^3/s (20-mgd) Plant
Determine Capital Cost
Use same approach as for 0.088-m^3/s (2-mgd) Plant.

Need Sludge Volume per Year:

104,000 gal/d x 365 days/yr = 38.0 x 10^6 gal/yr

Adjust Sludge Volume due to Loading Rate:

0.64 x 38.0 Mgal/yr = 24 Mgal/yr

From Figure 5-13 of the EPA Handbook (1) @ 3% solids:

Capital Cost = \$1,100,000

Determine O&M Cost
Use Figure 5-14 and actual sludge volume of 38 Mgal/yr.

Base Annual O&M Cost = \$140,000/yr

Determine Costs for Polymer Feed System
See approach used for 0.088-m^3/s (2-mgd) plant.

A.4 Vacuum Assisted Drying Beds

A.4.1 Design Examples

Design Example for a 0.088 m^3/s (2 mgd) Plant
Note: Since Vacuum Assisted Drying Beds (VADBs) are not typically used in 0.88-m^3/s (20-mgd) plants, a Design Example has been prepared only for the smaller 0.088-m^3/s (2-mgd) plant.

Sludge Quantity
Sludge Solids = 1,190 kg/d (2,600 lb/d)
Sludge Volume = 39,300 l/d (10,400 gpd)

Select a Solids Loading Rate
A typical upper limit solids loading rate for a VADB, without decanting any supernatant, is about 10 kg/m^2/cycle (2 lb/sq ft/cycle). However, one can assume that some decanting of supernatant can be accomplished and that a solids loading of 15 kg/m^2/cycle (3 lb/sq ft/cycle) can be used. A typical cycle can be completed in 24 hours or less.

It is desirable to have a minimum of two VADBs at a plant, and three are preferable for flexibility. Each bed should be sized to handle, at a minimum, 70 percent of an average daily sludge quantity.

Minimum Bed Size = (1,190 kg/d)(0.70)/15 kg/m^2/d
= 55 m^2 (590 sq ft)

Standard media plate sizes are 0.6 m x 0.6 m (2 ft x 2 ft) or 0.6 m x 1.2 m (2 ft x 4 ft).

Standard VADB sizes are 6.1 m x 6.1 m (20 ft x 20 ft) or 6.1 m x 12.2 m (20 ft x 40 ft). The area calculated above is in between the standard VADB sizes of 37 and 74 m^2 (400 and 800 sq ft).

Check bed solids loadings at 37 and 74 m^2 (400 and 800 sq ft):

(1,190 kg/d)/37 m^2 = 32 kg/m^2/cycle
(6.6 lb/sq ft/cycle)

(1,190 kg/d)/74 m^2 = 16 kg/m^2/cycle
(3.3 lb/sq ft/cycle)

The solids loading for a 37-m^2 (400-sq ft) bed would be too high to be acceptable as a design basis. The solids loading for a-74 m^2 (800-sq ft) bed would be very conservative if three beds are chosen. More than likely, if the 74-m^2 (800-sq ft) bed was chosen, only two beds would be designed.

A better design is to select three beds of about 56 m^2 (600 sq ft) each. This design allows more flexibility than with two overly large beds. With three beds, the dewatering system can operate an average of 5 days per week, yet it can handle a full 7 days' accumulation of sludge.

Size Polymer Feed System
From Section 6.4.3, the median polymer dosage at 13 VADB systems was approximately 10 g/kg (20 lb/ton) of solids. It is necessary to determine the volume of liquid emulsion type polymer to use. For example, a liquid emulsion type polymer may be 12 percent polymer by weight.

Thus, total weight of polymer is:

(10g/kg)/(0.12) = 83 g polymer/kg solids
(166 lb/ton)

If sludge is flowing to the VADB at a velocity of 0.6 m/s (2 ft/sec) through a 15 cm (6 in) pipe, the sludge feed rate is:

Sludge Flow
= Velocity x Cross-sectional Area
= (0.6 m/s)(π/4)(0.15 m)^2(1,000 l/m^3)(60 s/min)
= 636 l/min (168 gpm)

Each sludge flow volume of 636 l/min (168 gpm) has the following quantity of sludge solids:

(636 l/min)(1.0kg/l)(1.01)(0.03) = 19 kg/min
(42 lb/min)

Polymer Requirement
= (83 g polymer/kg solids)(19 kg/min)/1,000 g/kg
= 1.58 kg/min (3.48 lb/min)

If polymer is diluted 5 volumes water per 1 volume polymer:

Feed Volume = (6)(1.58 kg/min)/(1.0 kg/l)
= 9.5 l/min (2.5 gpm)

These calculations indicate that the polymer system should be capable of feeding a minimum of 7.5 to 75 l/min (2 to 20 gpm) of polymer feed solution and 1.25 to 12.5 kg/min (2.75 to 27.5 lb/min) of polymer.

A.4.2 Cost Analysis

The EPA Handbook (1) does not contain costs for a vacuum assisted drying bed. See Section 6.4.5 of this Design Manual for cost information.

A.5 Belt Filter Presses

A.5.1 Design Examples

Design Example for a 0.088-m^3/s (2-mgd) Plant

Sludge Quantity

Sludge Solids = 1,190 kg/d (2,600 lb/day)
Sludge Volume = 39,300 l/d (10,400 gpd)

Determine Size and Number of Belt Presses

At a plant of this size, one can assume that sludge dewatering operations would be restricted to 5 days per week and 8 hrs per day.

(1,190 kg/d)(2,600 lb/d)(7 d/wk)
= 8,330 kg/wk (18,200 lb/wk)

(39,300 l/d)(10,400 gpd)(7 d/wk)
= 275,100 l/wk (72,800 gal)/wk

From Section 7.2.6, a typical sludge throughput is 2.5 l/s (40 gpm) per meter belt width.

Hours Required Per Week
= (275,100 l/wk)/(2.5 l/s)(3,600 s/hr)
= 30.2 hr/wk/m of belt width

With 5 days per week and 8 hrs per day, there are 40 hrs per week available. 30.2 hrs is less than 40 hrs, therefore one 1-m belt press would provide enough capacity @ 6.0 hr/d.

Operation would be for 5 days/wk. One belt press would typically be sufficient for a plant of this size, provided there is some other backup means of dewatering or storing sludge for several weeks at most.

Size Polymer Feed System

Typical polymer requirements (from Section 7.2) are 2-8 g/kg (3-15 lb/ton).

Sludge Solids = (8,330 kg/wk)/(30.2 hr/wk)
= 276 kg/hr (608 lb/hr)

Design to provide sufficient capacity to feed up to 10 g/kg (20 lb/ton) polymer.

Polymer Requirements
= (276 kg/hr)(10 g polymer/kg)
= 2,760 g polymer/hr (6.2 lb/hr)

If polymer solution is @ 0.1 percent:

Feed Volume
= (2,760 g/hr) (0.001)(1 kg/l)(1000 g/kg)
= 2,760gl/hr (729 gal/hr)

Thus, the Polymer Feed system should be capable of feeding 3,000 g/hr (6.6 lb/hr) of dry polymer and 3,000 l/hr (790 gal/hr) of polymer solution.

Volume of Sludge Cake Produced

Determine Sludge Cake Specific Gravity (SSG). Assume the sludge cake is 20 percent solids and that the Digested Sludge Solids have a specific gravity of 1.4.

SSG [Equation 2-3 from (1)]
= 1/[[(20)/(100)(1.4)] + [(100 - 20)/100]]
= 1.06

Sludge Volume [Equation 2-1 from (1)]
= (276 kg/hr)(6.0 hr/d)/(1.06)(1.0)(0.20)
= 7,800 l/d (275 cu ft/d)
= 7.8 m^3/d (10 cu yd/d)

Belt Press Filtrate

Washwater requirements (vary with belt press manufacturer):

Assume 3.16 l/s (50 gpm)/m belt width (per Section 7.2).

Solids Capture: Assume 85% (worst case).

Solids in Filtrate: 276 kg/hr x (1 - 0.85) = 41 kg/hr (90 lb/hr)

Filtrate Flow Rate = Washwater Flow Rate
+ Digested Sludge Volume
- Dewatered Sludge Volume

= (3.16 l/s)(3,600 s/hr)
+ (275,100 l/wk)/(30.2 hr/wk)
- (7,800 l/d)/(6 hr/d)

= (11,400 + 9,100 - 1,300) l/hr
= 19,200 l/hr

Filtrate Volume Per Day
= (19,200 l/hr) (6 hr/d)
= 115,200 l/d (30,432 gpd)

Filtrate Solids Concentration
= (100)(41 kg/hr)/(19,200 l/hr)(1.0 kg/l)
= 0.21 percent solids
= 2,100 mg/l

Design Example for a 0.88-m^3/s (20-mgd) Plant
Sludge Quantity
Sludge Solids = 11,900 kg/d (26,000 lb/d)
Sludge Volume = 393,000 l/d (104,000 gpd)

Determine Size and Number of Belt Presses
Assume sludge dewatering operations are 16 hr/d and 7 days/wk. A typical sludge throughput is 2.5 l/s (40 gpm)/m belt width.

(2.5 l/s/m)(16 hr/d)(3600 s/hr)
= 146,000 l/d/m belt width required operating at 16 hr/d

(393,000 l/d)/(146,000 l/d/m)
= 2.7 m of belt required

At first glance, it appears that two 1.5-m belt presses could handle the sludge volume. However, if one belt press is out of service, could the other unit handle the total plant sludge if operated 24 hours a day?

Check: 146,000 l/dy/m x 1.5m x (24 hr/16 hr)
= 329,000 l/d, <393,000 l/d sludge flow

Thus, two 1.5-m belt presses are not sufficient.

Try two 2-m belt presses, Again, if one unit is out of service, the other unit must process the total sludge volume operating 24 hr/d.

Sludge Throughput in 24 hr
= (2.531 l/s/m)(24 hr/d)(3,600 s/hr)
= 219,000 l/d/m

(393,000 l/d)/(219,000 l/d/m)
= 1.8 m of belt required

One machine could temporarily handle the total sludge flow. Check the operating time for two 2-m belt presses:

Hours Required Per Day
= (393,000 l/d)(4 m belt)/(146,000 l/d/m)
= 10.8 hr/d

Polymer Feed System
See approach for (0.88 m^3/s) 2 mgd plant.

Volume of Sludge Cake Produced
Sludge Solids Feed Rate
= (11,900 kg/d)/(10.8 hr/d)
= 1,100 kg/hr (2,430 lb/hr)

Sludge Volume
= (1,100 kg/hr)(10.8 hr/d)/(1.06)(1.0 kg/l)(0.20)
= 56,000 l/d (14,800 gpd)
= 56 m^3/d (73 cu yd/d)

Belt Press Filtrate
Assume solids capture 85% (worst case).

Solids in Filtrate: 1,100 kg/hr x (1 - 0.85)
= 165 kg/hr (364 lb/hr)

Filtrate Flow Rate = Washwater Flow Rate
+ Digested Sludge Volume
- Dewatered Sludge Volume

= (3.16 l/s)(3,600 s/hr)
+ (393,000 l/wk)/(10.8 hr/wk)
- (56,000 l/d)/(10.8 hr/d)

= (45,000 + 36,400 - 5,200) l/hr
= 76,600 l/hr

Filtrate Volume per Day
= (76,700 l/hr)(10.8 hr/d)
= 828,000 l/d (219,000 gpd)

Filtrate Solids Concentration
= (100)(165 kg/hr)(76,700 l/hr)(1.0 kg/l)
= 0.22 percent solids
= 2,200 mg/l

A.5.2 Cost Analysis

Cost Analysis for a 0.088-m^3/s (2-mgd) Plant
Determine Capital Cost
Use Figure 5-4, Base Capital Cost of Belt Filter Press, page 75 of EPA Handbook (1).

Need Sludge Volume per year (from Design Example):

(10,400 gpd)(365 days/yr) = 3.80 x 10^6 gal/yr

Figure 5-4 is based upon a solids loading rate of 500 lb/hr per meter for 2 percent solids and 650 lb/hr per meter for 4 percent solids, or about 575 lb/hr/per meter for 3 percent solids.

Check Solids Loading in Design Example:

(40 gal/min-m)(60 min/hr)(8.34 lb/gal)(1.01)(3/100)
= 606 lb/hr/ m

606 ≈ 575 lb/hr. No need to adjust curve for this difference.

Figure 5-4 is based upon 8 hr/day, 7 days/wk operation.

8 hr/d x 7 d/wk = 56 hr/wk operation

From Figure 5-4 @ 3% solids and 3.8 Mgal/yr:

Capital Cost = $270,000

From Design Example:

With one 1-m belt press, a 40 hr/wk operation (including startup, shutdown, and cleanup) is common. With fewer hours of operation, the belt press would need to be larger. Therefore, the capital cost obtained from the curve is too low.

Adjust sludge volume by ratio of 56/40 = 1.40

1.40 x 3.8 Mgal/yr = 5.3 Mgal/yr

From Figure 5-4 @ 3% solids and 5.3 Mgal/yr:

Capital Cost = $290,000.

Note: The adjustment procedure for different hours of operation is also illustrated on page 27 of the EPA Handbook (1).

Determine O&M Costs
Use Figure 5-5 of the EPA Handbook, which is for Base Annual O&M.

For the actual sludge flow of 3.8 Mgal/yr and 3% solids:

Base Annual O&M Cost = $0.012 million/yr
= $12,000/yr

This is based upon a labor cost of $13.50/hr and an electricity cost of $0.094/kWh. Using the same procedure and Figure 5-6, the labor man hours, material costs, and electrical energy costs in kWh/yr can be obtained.

Determine Costs for Polymer Feed System
Capital and O&M costs for the polymer feed system can be computed using the EPA Handbook (1), as shown in Section A.3.2 of this manual.

Cost Analysis for a 0.88 m^3/s (20 mgd) Plant
Determine Capital Cost
Use Figure 5-4 of the EPA Handbook (1).

Sludge Volume per year = 38 x 10^6 gal/yr

Figure 5-4 is based on 8 hr/day, 7 d/wk operation. From the design example, operation is 10.8 hr/d, 7 d/wk. Thus, the belt presses could be smaller (the capital cost would be too high).

Adjust Sludge Volume by ratio of 8/10.8 = 0.74

0.74 x 38 Mgal/yr = 28 Mgal/yr

From Figure 5-4 @ 3% solids and 28 Mgal/yr,

Capital Cost = $650,000.

Determine O&M Costs
From Figure 5-5 @ 3% solids and 38 Mgal/yr,

O&M Cost = $75,000/yr.

Determine Costs for Polymer Feed System
Use procedure shown in Section A.3.2 of this manual to obtain costs from the EPA Handbook (1).

A.6 Solid Bowl Centrifuges

A.6.1 Design Examples

Design Example for a 0.088-m3/s (2-mgd) plant
Sludge Quantity
Sludge Solids = 1,190 kg/d (2,600 lb/d)
Sludge Volume = 39,300 l/d (10,400 gpd)
Sludge Solids Concentration = 3%

Determine Size and Number of Centrifuges
Refer to section 7.3 of this manual. Table 7-10 shows suggested capacities and numbers of centrifuges for various plant sizes. For the 0.088-m^3/s (2-mgd) plant, the sludge flow is 40 m^3/d, which is essentially equal to the 39,300 l/d used in this example. The number of hours of operation, seven, is reasonable as is the number of centrifuges (one duty, one standby). It is assumed that centrifuge operations are 7 hr/d, 7 d/wk.

Write Performance Specification
As described in Section 7.3, for a specific sludge flow rate, the actual size (both diameter and length) of the required centrifuge varies from supplier to supplier. The low speed centrifuge compared to the high-speed centrifuge of similar capacity will be larger in diameter.

The typical engineering design of a centrifuge dewatering facility will select the number of units and will specify the performance which must be achieved. Typically, the design engineer does not select the actual model and size of centrifuge required, unless on-site, side-by-side testing of several different centrifuge models has been conducted.

A typical performance specification includes both design sludge characteristics and performance requirements. Two such sections from a complete specification on a centrifuge are presented for the 0.088-m^3/s (2-mgd) plant as follows:

<u>Job Conditions</u>
A. Design Sludge Characteristics

1. Mixed anaerobically digested sludge to centrifuge
   a. Total solids
      1) Average: 3%
      2) Range: 2.5-3.5%
   b. Volatile solids
      1) Average: 54% of total solids
      2) Range: 50-70% of total solids
   c. Ratio WAS to primary sludge before digestion
      1) Average: 38:62
      2) Range: 35:65 to 50:50
   d. Average temperature: 21°C

Design and Performance

A. The following performance requirements shall be met for the sludge characteristics specified herein.

1. Digested sludge dewatering

   a. Cake solids: 18% minimum

   b. Recovery of suspended solids: 90% minimum

   c. Polyelectrolyte allowed to achieve required cake solids and recovery: 4 g polymer/kg dry sludge solids max. (8 lb polymer/ton max.)

   d. Design flow rate: 5.7 m^3/hr (25 gpm)

A complete engineering specification for solid-bowl centrifuges contains many more requirements than merely the performance specification. Other items that would be included are descriptions of required submittals of shop drawings and product data, equipment, O&M manuals, product delivery requirements, and guarantees. Specifications would also include material requirements for bowls, conveyors, and for the type of abrasion resistant materials to be used. Also included are requirements for the fabrication and manufacture of the many components of the centrifuge including the motor and drive, the type of backdrive system, and the electrical controls.

Polymer Feed System
Although the performance specification requires the centrifuge to use a maximum polymer dosage of 4 g/kg (8 lb/ton), the polymer system should be designed to be able to deliver a minimum of 10 g/kg (20 lb/ton).

Volume of Sludge Cake Produced (Assume 18 Percent Solids):

Determine Sludge Cake Specific Gravity, SSG
SSG [Equation 2-3 from (1)]
= 1/[[(18)/(100)(1.4)] + [(100 -18)/100]]
= 1.05

Sludge Volume [Equation 2-1 from (1)]
= (1,190 kg/d)/(1.05)(1.0)(0.18)
= 6,300 l/d (1,700 gpd)
= 6.3 m^3/d (8.2 cu yd/d)

Centrate
Centrate Volume = Digested Sludge Volume - Dewatered Sludge Volume

= 39,300 l/d - 6,300 l/d
= 33,000 l/d (8,700 gpd)

Assume 90% solids capture (worst case):

Solids in Centrate: 1,190 kg/d (1 - 0.90) = 119 kg/d (262 lb/d)

Centrate Solids Concentration
= (100)(119 kg/d)/(33,000 l/d)(1.0 kg/l)
= 0.36 percent solids
= 3,600 mg/l

Design Example for a 0.88-m^3/s (20-mgd) Plant
Sludge Quantity
Sludge Solids = 11,900 kg/d (26,000 lb/d)
Sludge Volume = 393,000 l/d (104,000 gpd)

Determine Size and Number of Centrifuges
Refer to Section 7.3.

For a 0.88-m^3/s (20-mgd) plant size, the sludge flow is 320 m^3/d, which is 19 percent lower than the sludge volume assumed in these design examples. The number of hours of operation (15) is too low for 12-m^3/hr (53-gpm) centrifuges. This design should select three 13-m^3/hr (58-gpm) centrifuges, based on 15 hr/day operation. The 15 hr/day operation is desirable because it allows 2-shift operation and one hour for required cleanup.

Write Performance Specification
The performance specification would be identical to the one for the 0.088-m^3/s (2-mgd plant) with the exception that the design sludge flow rate is 13 m^3/hr (58 gpm).

A.6.2 Cost Analysis

Cost Analysis for a 0.088 m^3/s (2 mgd) Plant
Determine Capital Cost
Use Figure 5-1, Base Capital Cost of Centrifuge, of the EPA Handbook (1).

Sludge Volume per Year (from Section A.3.2)
= 3.80 x 10^6 gal/yr

Figure 5-1 is based upon 8 hr/d, 7 d/wk operation. In this example, only 7 hr/d operation is assumed. Thus, the centrifuges in the example need to be slightly larger than assumed in Figure 5-1.

Adjust annual sludge volume by ratio of 8/7 = 1.14

1.14 x 3.8 Mgal/yr = 4.3 Mgal/yr

From Figure 5-1 @ 3% solids:

Capital Cost = $260,000.

Determine O&M Cost
From Figure 5-2 and actual sludge volume of 3.8 Mgal/yr,

O&M Cost = $30,000/yr.

Cost Analysis for a 0.88 m^3/s (20 mgd) Plant
Determine Capital Cost
Use Figure 5-1, page 72 of EPA Handbook (1).

Sludge Volume per Year (from Section A.3.2)
= 38 Mgal/yr

Adjust annual sludge volume by ratio of 8/7 = 1.14

1.14 x 38 Mgal/yr = 43 Mgal/yr

From Figure 5-1, Capital Cost = $620,000

Determine O&M Cost
From Figure 5-2 and actual sludge volume of 38 Mgal/yr,

O&M Cost = $82,000/yr

Polymer Feed Systems
Figures 6-19 and 6-20 of the Handbook would be used to obtain polymer feed system capital costs as was done in section A.3.2. The design polymer dosage should be 10 lb/ton.

For O&M costs, Figures 6-22 and 6-23 would be used. The operating polymer dosage should be 6 lb/ton.

A.7 Filter Presses

A.7.1 Design Examples

Design Example for a 0.088 m^3/s (2 mgd) Plant
Sludge Quantity
Sludge Solids = 1,190 kg/d (2,600 lb/d)
Sludge Volume = 39,300 l/d (10,400 gpd)
Sludge Solids Concentration = 3%

At this size plant, assume 8 hr/d, 5 d/wk operations.

(1,190 kg/d)(7 d/wk) = 8,330 kg/wk
(18,200 lb)/wk

(39,300 l/d)(7 d/wk) = 275,100 l/wk
(72,800 gal)/week

Determine Dewatered Sludge Cake Volume
Assume 40 percent solids cake and that this is equivalent to a cake density of 1,140 kg/m^3 (71 lb/cu ft)

Cake Volume per Day of Operation
= (8,330 kg/wk)/(5 d/wk)(0.40)(1,140 kg/m^3)
= 3.65 m^3/d (129 cu ft/d)
= 3,650 l/d

Determine Size of Filter Press Required
Assume one press capable of dewatering all the sludge volume per day. Assume in an 8-hour day that, conservatively, 3 filter cycles could be completed.

Required Filter Volume = (3,650 l/d)(3)
= 1,220 l (43 cu ft)

Number of Filter Presses Required
For a wastewater treatment plant this small, only 0.088 m^3/s (2 mgd), a filter press would be selected for dewatering only if (1) a very dry sludge cake is required, and (2) very little land area is available. In this case, dewatering reliability is probably critical, and two filter presses should be provided. One filter press would serve as a standby unit.

Polymer Dosage
A draft EPA report on recessed plate filter press design and operation (2) describes conditioning with a highly charged cationic polymer to produce sludge cake of 35% solids. One facility reported reducing its chemical conditioning cost from $76/dry ton of dewatered sludge to $24/dry ton by using polymer instead of lime and ferric chloride. Determination of the size and costs for a polymer feed system can be conducted as done in Section A.3.2 of this manual.

Before a filter press installation is designed with only polymer conditioning, numerous on-site tests of the particular sludge and polymer must be conducted. Not all sludges can be dewatered on a filter press that uses only polymer conditioning.

Filtrate
A filter press can normally achieve a solids capture of 95 to 99% when conditioned with lime and ferric chloride. For polymer conditioning, a worst case capture of 90% can be assumed. Calculation of the filtrate quantity would be similar to the filtrate calculation for belt filter presses in Section A.5.1.

Design Example for a 0.88 m^3/s (20 mgd) Plant
Sludge Quantity
Sludge Solids = 11,900 kg/d (26,000 lb/d)
Sludge Volume = 393,000 l/d (104,000 gpd)
Sludge Solids Concentration = 3%

At this plant, assume 16 hr/d, 7 d/wk operation.

(11,900 kg/d)(7 d/wk) = 83,300 kg/wk
(182,000 lb/wk)

(393,000 l/d)(7 d/wk) = 2,751,000 l/wk
(728,000 gal/wk)

= 2,751 m^3/wk
(3,600 yd3/wk)

Determine Dewatered Sludge Cake Volume
Assume 40% Solids and 1,140 kg/m^3 (71 lb/cu ft)

Cake Volume per Day of Operation
= (8,330 kg/wk)/(7 d/wk) (0.40)(1,140 kg/m^3)
= 26.1 m3/d (922 cu ft/d)

Determine Size of Filter Press Required
If a typical cycle time is 2.5 hours, then during a 16 hour work day,

(16 hr/d)/(2.5 hr/cycle) = 6.4 cycles/d
(use 6 cycles/d)

Required Filter Volume
= (26.1 m^3/d)/(6 cycles/d)
= 4.3 m^3 (152 cu ft)

Number of Filter Presses Required
Two filter presses should be available for service for reliability. Two filter presses of 4.3-m^3 volume each should be installed, although the individual designer can vary the actual size selected.

One reason for choosing two presses is that it allows a reduction to 8 hr/d, 7 d/wk operation. Then, if one filter press is out of service, the 16 hr/d and 7 dy/wk operation could be used. Also, if the sludge proves more difficult to dewater than anticipated, there will be sufficient capacity available.

A.7.2 Cost Analyses

Cost Analysis for a 0.088 m^3/s (2 mgd) Plant
Determine Capital Cost
Use Figure 5-7, Base Capital Cost of Recessed Plate Filter Press, page 78 of the EPA Handbook (1).

Sludge Volume per Year = 3.80 x 10^6 gal/yr

Figure 5-7 is based on 8 hr/d, 7 d/wk operation. In this example, operation will be only 5 days per week.

Adjust annual sludge volume by ratio of 56/40 = 1.4

1.4 x 3.8 Mgal/yr = 5.3 Mgal/yr

From Figure 5-7 @ 3% solids:

Capital Cost = $300,000 for 1 unit

Appendix A-10 from the EPA Handbook (1) lists the background assumptions for the cost figures on the filter press. For total press chamber volumes below 450 ft3, only one filter press is included. Therefore, to get a cost for the two filter presses (one standby), the cost for one must be multiplied by 2. Some reduction in cost is possible, for example 15 percent, in construction of two identical units.

(1-0.15) x $600,000 = $510,000

Determine O&M Cost
From Figure 5-8 and actual sludge volume of 3.8 Mgal/yr,

O&M Cost = $15,000/yr

Cost Analysis for a 0.88 m^3/s (20 mgd) Plant
Determine Capital Cost
Use Figure 5-7 of the EPA Handbook (1).

Sludge Volume per Year = 38 x 10^6 gal/yr

Figure 5-7 is based on 8 hr/day, 7 days per week operation. In this example, operation will be 16 hr/day, 7 days per week.

Adjust annual sludge volume by 8/16 = 0.50

0.50 x 38 Mgal/yr = 19 Mgal/yr

From Figure 5-7 @ 3% solids,

Capital Cost = $520,000

For this design, two 4.3-m^3 (152-cu ft) presses, the cost would be calculated as for the 0.088-m^3/s plant:

(1-0.15) x $1,040,000 = $880,000

Determine O&M Cost
From Figure 5-8 and actual sludge volume of 38 Mgal/yr,

O&M Cost = $58,000/yr

A.8 References

1. *Handbook: Estimating Sludge Management Costs*. EPA-625/6-85/010, United States Environmental Protection Agency, Center for Environmental Research Information, Cincinnati, OH, p. 540, 1985.

2. *Recessed Plate Filter Presses* (Design Information Report). EPA-600/M-86/017, United States Environmental Protection Agency, Center for Environmental Research Information, Cincinnati, OH, 1986.

Appendix B

Operation and Maintenance, Mechanical Dewatering Systems

B.1 Introduction

This appendix presents operation and maintenance information for several mechanical dewatering systems. Sections have been included for belt filter presses, centrifuges, filter presses, and vacuum filters. Sample log sheets can be used in the development of a record-keeping program.

B.2 Belt Filter Presses

Replacement of the filter belts is one of the most common maintenance items. The main reasons for failure of the belts are tearing at the clipper seam, inferior quality belt material, ineffective tracking systems and/or poor operation and maintenance. Clipper seam failure usually occurs because of inferior construction of the seam or sharp edges on the doctor blades. The seams should be epoxy coated on each side and doctor blades should be made of polyethylene instead of metal. Furthermore, ineffective tracking systems can cause the edges of the belt to wear and fray. Poorly designed and maintained high pressure washwater pumps and spray nozzles can also cause belt wear. The pump may not deliver sufficient pressure and flow to the spray nozzles, which in turn may clog. Without an even flow of water across the belt, it will blind. Blinding leads to belt wrinkling and creasing, thus reducing belt life. Because of improvements in belt and press design, especially in the area of clipper seam construction and belt tracking and tensioning systems, belt life has increased tremendously. Average belt life is about 2,700 running hours with a range of 400-12,000 running hours (1). Table B-1 (2) shows typical causes of belt wear problems and possible solutions.

Belt filter presses are probably the most energy conservative and therefore, the most economical mechanical dewatering units to operate, since they have very low power requirements. The average requirement is about 5.7 kW (8 hp) per meter belt width, which is considerably lower than other types of mechanical dewatering equipment. Some models are as low as 0.8 kW (1 hp) per meter belt width (3).

Staffing requirements are also low. Table B-2 (3) summarizes typical operator requirements.

Table B-1. Causes and Prevention of Belt Wear

Cause	Preventive Measure
1. Inferior belt material and inaccurate dimensions	Purchase only high-quality belts with a guaranteed life from well-known manufacturers.
2. Wear at clipper seam	Purchase belts with highly durable seams, especially those epoxy coated on both sides. Check doctor blades to ensure that there are no sharp edges on which the clipper can get caught and tear.
3. Misalignment of rollers	Check and adjust roller alignment. Ensure well-functioning tensioning/tracking systems.
4. Belt shifting or creasing	Ensure that sludge is distributed evenly across the belt. Ensure that belt washing system is adequate.

Table B-2. Number of Operators Required Per Belt Filter Press

Number of Units	Number of Operators Per Unit
1	1.09
2	0.66
3	0.36
4 or more	0.33

In order for a belt filter press or for that matter, any other type of mechanical dewatering unit to function effectively, it must be compatible with both upstream and downstream processes. Often, operational problems can be reduced simply by specifying the appropriate equipment and by properly training the operators of the equipment. Table B-3 (2) lists typical causes and preventive measures to reduce operational problems associated with belt filter presses.

Process control is extremely important to ensure optimum performance of the dewatering system. It is important for the operator to keep records of all press performance parameters. A typical operational log sheet is shown in Figure B-1. The operator can determine how well the press is performing from the information provided by this log. Sample points must be located at various places throughout the system. At least once per shift, a sample should be taken of

Table B-3. Causes and Prevention of Operational Problems of Belt Filter Presses

Cause	Preventive Measure
1. Improperly conditioned sludge due to:	1. Improve conditioning practices by:
a. Varying characteristics of sludge feed	a. Assuring continuity of sludge through blending of sludge prior to dewatering
b. Improper chemical selection or dosage rate	b. Polymers should be carefully selected and tested. Selection and dosage rate should be checked frequently, particularly when changes in sludge characteristics are expected
c. Improper point of application.	c. The point at which polymer is applied should be reviewed and revised.
2. Insufficient gravity drainage of sludge	2. Evaluate: a. Press speed/drainage time b. Polymer selection and sludge conditoning system.
3. Loss of sludge from between belts	3. Reduce belt tension
4. Poor housekeeping	4. Train operators to properly maintain the press area. Provide steam cleaning equipment to assist cleanup.
5. Poor safety practices, including:	5. Provide safety training and stringently enforce rules to keep safety equipment in place.
a. Removal of spray and other equipment guards to facilitate operation	a. Design guards for ease of replacement after removal
b. Deactivation of trip-wire and limit switches to facilitate access to unit.	b. Design safety equipment to minimize interference with operation.

the feed sludge to the press, cake discharge, and filtrate. However, composite sampling at these three locations is the best choice. A composite sample will give a better picture of press performance. Total solids should be determined on the feed and cake samples and total suspended solids should be determined for the filtrate. *Standard Methods for the Examination of Water and Wastewater* (4) should be used as the source for all laboratory protocol. The press operator should be required to make the following calculations:

$$\text{Throughput, kg/hr} = Q\ C \times 1{,}000$$

where,

Q = flow to press in m^3/hr
C = concentration of solids in feed sludge in mg/l

$$\text{Throughput, lb/hr} = 8.345\ Q\ C$$

where,

Q = flow to press in gal/hr
C = chemical fraction of total solids in feed sludge

Capture, %

$$= 100 \times [(\%\text{Cake Solids})(\%\text{Feed}-\%\text{Filtrate})] \div [(\%\text{Feed Solids})(\%\text{Cake}-\%\text{Filtrate})]$$

B.3 Solid Bowl Centrifuges

In the operation of centrifuges, the performance criteria of high TSS recovery, driest cake solids and high sludge capacity are not mutually compatible. To achieve the driest cake, the centrifuge solids capacity may be limited and higher polymer dosages may be required. While a lower solids recovery will increase cake solids, this is not an acceptable long-term operating mode. The fines lost to the centrate will eventually be recycled until removed from the system or lost, thus impairing the final effluent.

The general effect of process variables on the cake solids and recovery is shown in Table B-4. While temperature is not a common variable available to the operator, heating of the feed sludge can result in a significant increase in cake concentration. In some cases, polymer addition result sin increased moisture content because of the capture of fines; in other cases, moisture content is decreased due to higher structural strength and better liquid expression from the cake.

Table B-5 shows the machine variables that can affect solids recovery and cake solids. While higher bowl speeds can produce a drier cake, at some point the g forces will be so high that the cake will not convey out of the pool. Deepening the pool under these conditions may alleviate the conveying problems, but the cake will be wetter. However, reducing the differential speed can effect a solids "dam" in the centrifuge and thereby still maintain high solids, with the ability to convey the solids. Generally, weak structured solids, like alum and activated sludge, often respond best to lower gravitational forces. Reducing the differential speed will produce a drier cake, but at some point it will impair solids recovery due to the depth of sludge in the pool.

Often, it is possible to produce a drier sludge cake by increasing the polymer dosage. This may be cost effective for a combustion operation. For example, if the cake solids were increased from 20 to 24 percent TS with $11.00/Mg dry solids ($10.00/ton) additional

TYPE OF SLUDGE: ______________________ PRESS NUMBER: ____________

DATE TIME	TEST NO.	FEED			POLYMER				FILTRATE		RECOVERY, %	CAKE DRY SOLIDS, %	HP/PSI	SCREEN TENSION	WASH WATER, PSI
		GPM	TOTAL SOLIDS, %	LBS/HR	TYPE	ML/MIN	PPM	LBS/D.T.	GPM	PPM					

REMARKS:

Figure B-1. Sample log sheet for belt filter press.

Table B-4. **Effect of Process Variables on Recovery and Cake Solids***

Process Variables	Feed Rate	Feed Consistency	Temperature	Flocculants
To improve recovery	-	+	+	+
To improve cake solids	-	-	+	-, +

* The process variables are highly interactive and any evaluation of a single parameter requires maintaining the other process variables constant. Machine variables should not by varied until final optimization.
- sign: reduction in the process variable produces the desired effect.
+ sign: increase in the process variable produces the desired effect.

Table B-5. **Effect of Machine Variables on Recovery and Cake Solids***

Machine Variables	Bowl Rate	Pool Volume	Conveyor Speed
To improve recovery	-, +	+	-
To improve cake solids	-, +	-	-

* In each case the TSS recovery is considered to be about 90 percent for comparative evaluation and other factors are held constant, i.e., feed rate, feed solids, etc. If cake structural characteristics are weak, cake solids and recovery may decrease due to solids slipping back down the beach slope. In that case it is necessary to increase polymer dosage to maintain recovery.
- sign: reduction in the machine variable produces the desired effect.
+ sign: increase in the machine variable produces the desired effect.

polymer cost, the fuel used by the furnace could be reduced 165-210 l/Mg dry solids (40-50 gal/ton), saving about $30.00/Mg dry solids ($27.00/ton) in fuel costs. Further, the furnace solids capacity would increase 30-40 percent. The operator should periodically test this condition.

High polymer usage can be due to a number of factors. Full-scale and laboratory testing can be conducted to help identify the reasons for the high polymer dosage. Operators should periodically evaluate new polymers as sludge properties change and new products become available. The procedure suggested for evaluating high polymer dosage and the resulting cake solids concentration is outlined below.

There are a number of reasons for excessive polymer usage. These causes may be evaluated step-by-step:

1. The polymer is not effective for the specific sludge.
Only a series of polymer trials can determine the best polymer. Lab tests can generally be used to sort out the best two or three. The sludge should appear granular when it is poured two or three times from beaker to beaker.

2. The polymer is not fully into solution and only partially effective.
The polymer should be tested at the normal age employed, then allow 6-8 hours more before retesting. The higher the aging concentration, the slower the polymer molecules go into solution and unravels. Use lab tests at the normal age solution concentration and also at half concentration. Test sludge flocculation at 4 hours and 12 hours.

3. There is a charge on the sludge that must be neutralized first.
Mix 1 kg/Mg dry solids (2 lb/ton) of anionic polymer (dry basis) with 2-3 liters of sludge, so there is enough treated sludge for several cationic polymer tests. Wait 10 minutes after mixing in the anionic polymer, then add increasing dosages (e.g., 6-12-18 lb/ton or 3-6-9 kg/Mg) of cationic polymer to each sludge mixture in turn. Repeat the process with another batch of sludge, using 2 kg/Mg dry solids (4 lb/ton) of anionic polymer. Visually compare the floc structure of these mixtures to a chemically dosed mixture without anionic polymer. The matrix of these mixtures is shown below.

		Anionic Polymer Dosage (lb/ton)		
Cationic Polymer Dosage (lb/ton)		0	2	4
	6	X_1	X	X
	12	X	X_2	X
	18	X	X	X_3

Record visual observations using 200 ml sludge in 300 ml beakers and pouring it back and forth three times. Good polymer treatment will result in solids granulation and free water. Run tests marked X_1, X_2, and X_3 first to determine where to concentrate testing.

It is possible that 10-20 kg/Mg dry solids (20-40 lb/ton) $FeCl_3$ will be effective in place of anionic polymer. It may be a lower cost alternative to anionic polymer. Corrosive properties of $FeCl_3$ must be kept in mind.

4. The polymer and sludge are not effectively mixed at the introduction point.
If the full-scale results are the same as the lab tests, mixing is generally not the problem. When lab tests indicate lower dosage, inadequate mixing is a possible cause. Breakup when mixing can be due to a weak floc or shear, in which case another polymer should be considered.

5. The centrifuge is overloaded on a solids basis, which is impairing the clarification efficiency.
Solids overloading can result in high polymer dosage. If the scroll differential speed is increased and the

centrate clears, this indicates that the centrifuge was exceeding its Beta capacity (solids discharge capacity) and thus affecting clarification. This is especially true when the objective is to produce the maximum cake solids. If the cake becomes too wet at the higher differential speed, the feed rate must be reduced and then the differential speed to maximize cake solids.

The evaluation procedures 1 through 4 above will isolate problems associated with polymer selection, sludge characteristics, and polymer handling. Once the polymer and feeding have been optimized, the best polymer can be evaluated, using Step 5 to determine whether a limiting solids capacity is affecting polymer cost and performance.

The following tests to determine solids capacity are recommended:

A. Set up centrifuge to operate at a specific maximum torque setting, auto or manually controlled, which will be considered 100 percent capacity.

B. Feed centrifuge 25, 50, 75 and 100 percent of solids capacity.

C. Adjust polymer dosage for each flow to about the same effluent clarity-slightly dirty (90-95 percent recovery). Use lab centrifuge spin tests for accuracy and quick analysis of the probable recovery.

D. Allow 15 minutes for stabilization and then take three samples of feed, cake, and centrate at 15, 20 and 25 minutes (after machine conditions have been established) for composite.

E. Measure:

Feed rate
Feed TSS-keep the same, if possible
Cake TS
Centrate TSS-keep the same, as much as possible
Polymer rate
Differential speed

F. Plot:

1. Cake solids rate (lb/hr or kg/hr) vs. cake solids (% TS)

2. Cake solids (lb/hr or kg/hr) vs. polymer dosage (lb/hr or kg/hr)

3. Recovery (% TSS) vs. feed rate (lb TSS/hr and gpm or kg TSS and l/s) -- should be constant if polymer is adjusted to same clarity.

The cake solids (percent TS) will normally decrease slightly with increasing solids rate (lb/hr or kg/hr). If the differential speed is increasing significantly during the test in order to maintain torque within limits and centrate quality, then the cake may be significantly wetter at higher feed solids rate (lb/hr or kg/hr). Polymer dosage may also increase to help maintain solids volume in the centrifuge. If the polymer dosage is increasing and the cake becoming wetter, then the solids capacity of the centrifuge has probably been exceeded.

If there is a significant difference in the TSS recovery, this can cause the results to be misleading. Recovery should be maintained in the range of 90-95 percent of the feed solids.

To fully evaluate the effect of auto torque control, conduct the following tests. Based on the earlier test, use a feed rate that indicates that the unit is within +10 percent of its full capacity. Operate the differential speed manually at 25, 50, 75 and 100 percent of full torque.

A. Adjust chemical dosage as necessary.

B. Adjust eddy current or auto torque backdrive as necessary.

C. Operate for 15 minutes for stabilization and sample at 15, 20 and 25 minutes for composite.

D. Determine:

1. Differential speed, rpm

2. Feed rate, gpm and lb TSS/hr or l/s and kg TSS/hr

3. Polymer rate, gpm and lb/ton dry solids or l/s and kg/Mg dry solids

E. Plot:

1. Differential speed (rpm) vs. cake solids (% TSS)

2. Differential speed (rpm) vs. polymer (lb/ton or kg/Mg)

3. Polymer dosage vs. cake solids (%) and cake rate (lb TS/hr or kg TS/hr)

This test series should be run on the same feed concentration and recovery as much as possible so that solids discharge rate does not affect results. These results will define the optimum operation and determine the benefits available from the addition of an automatic torque-controlled backdrive. These optimization tests should be run periodically, even when the machine has an auto backdrive.

B.3.1 Record-keeping
Good record-keeping helps to detect long-term changes in the performance of the centrifuge. Performance differences could be due to the wear of the scroll, changes in sludge characteristics that require changes in bowl speed, improper pond setting, inappropriate bowl/scroll differential speed, etc. Performance fluctuations should lead to an optimization study of the centrifuge. As indicated earlier, polymer should also be continually evaluated. Communication with other centrifuge operators will also help to keep abreast of new developments.

Data log sheets should be maintained at the centrifuge and periodically checked to determine changes in the performance characteristics of cake solids, recovery, solids capacity, and polymer dosage.

Suggested data log sheets for daily and monthly operation of the centrifuge are shown in Figures B-2 and B-3. Representative samples should be collected and analyzed daily. This data, upon review could provide the basic insight necessary to optimize the operation.

On a regular basis, long-term trend lines should be plotted for the following parameters:

Feed Rate, lb TSS/hr or kg TSS/hr
Solids Concentration, % TS
Polymer Dosage, lb/ton TSS or kg/Mg TSS
Solids Recovery, %

An example is shown in Figure B-4. This trend log should be posted and updated weekly in an area visible to the operators.

As with any piece of machinery, the maintenance program is an essential part of a successful operation. Observations, such as an increase in vibration, could lead to a planned overhaul, thus avoiding a damaging breakdown. Reviewing and understanding the manufacturer's O&M manual will help lower O&M costs, reducing downtime and improving performance.

B.4 Filter Presses
Filter presses operate in a discontinuous or fill-and-dump mode. Since this mode of operation varies substantially with time, it imposes some special requirements when evaluating performance and developing effective operating patterns.

For example, it is not possible to take grab samples at any time during the operation on the dewatered or cake side of the operation and obtain any genuinely valuable information about performance. As a consequence, to obtain a full description of machine operation, a specific, directed sampling program must be created. Ideally, one would want to determine the volume of filtrate, which could be used to confirm the cake solids for any particular solids and total volume input to the machine.

B.4.1 The Sludge Conditioning System
At the onset of any performance evaluation or prior to startup, one must recognize the process key that controls all performance evaluations dealing with domestic wastewater sludges is proper sludge conditioning.

Many sludge conditioning problems do not derive directly from planned conditioning techniques but from failure of the ancillary or support equipment. This is particularly true when using ferric chloride (or sulfate) and lime to condition sludges, although polymer conditioning systems can and do have major maintenance problems. The purpose of the sludge conditioning system is to promote water release, increase cake solids, to permit a high solids recovery (generally greater than 98 percent) and to enhance cake removal. The conditioning system may involve the feeding of iron or aluminum salts, lime, and/or polymer; the delivery of a precoat material such as diatomaceous earth or incinerator ash; and/or or the addition of a body feed. All or any one or more of these materials may be employed on a particular sludge.

The management of the lime feed storage and preparation system is tedious, difficult, and demanding. Yet, when lime is employed, it is absolutely essential that it be available at the mixing tank. The operator should determine the availability of lime and the functioning capability of the lime feed system on a routine basis. Designers are urged to assure the easy maintenance and operation of lime receiving, storage, slaking, dissolution, and delivery systems. The very nature of hydrated lime slurries requires special attention. The engineer should design a system that the operator can rely on to function properly.

Too little lime means the sludge usually is improperly conditioned. Too much lime increases the rate at which filter cloths and plates become fouled with calcium hydroxide and calcium carbonate, that is, scale. Thus, excessive lime conditioning increases the frequency of acid washing, leading to increased downtime.

Sludge aging should be minimized, as should be the storage of sludge for any substantial period between the wet-end and the solids handling train or in the sludge handling train. Varying the holding time, for example, from one day to three days, will adversely effect the ability to condition the sludge and will increase the chemical requirements, particularly of polymers but also of iron salts and lime. Ferric chloride and lime are usually selected as the conditioning chemicals when complete control over

DAILY CENTRIFUGE OPERATOR LOG

DATE: ________

TIME	MACH. NO.	FEED RATE		% TSS	CAKE	POLY FEED		DILUTION	SAMPLES	TORQUE	DIFF SPEED	COMMENTS
		GPM	% TSS	ml/50 ml	% TSS	SETTING gpm		H_2O gpm	TAKEN	FT-LB	rpm	
2:00												
4:00												
6:00												
8:00												
10:00												
12:00												
14:00												
16:00												
18:00												
20:00												
22:00												
24:00												

FEED TOTALIZER	POLYMER TOTALIZER	POLYMER TYPE	OPERATORS
END ________	END ________	________	1. ________
START ________	START ________	________	2. ________
TOTAL ________	TOTAL ________	________	3. ________

Figure B-2. Sample daily log sheet for centrifuge.

Figure B-3. Sample monthly log sheet for centrifuge.

MONTHLY CENTRIFUGE PERFORMANCE LOG

MONTH: ______ YEAR: ______ PLANT: ______________

DATE	MACH. NUMBER	FEED SOLIDS GPM	FEED SOLIDS % TSS	FEED SOLIDS LB/HR	O'FLOW % TSS	CAKE % TSS	CAKE LB/HR	RECOVERY % TSS	POLYMER LB/TON DS	OPERATION HRS/DAY	COMMENTS
1											
2											
3											
4											
5											
6											
7											
8											
9											
10											
11											
12											
13											
14											
15											
16											
17											
18											
19											
20											
21											
22											
23											
24											
25											
26											
27											
28											
29											
30											
31											
TOTAL											
MIN.											
AVG											
MAX											

CENTRIFUGE TYPE: ______ BOWL SPEED: ______ RPM SLUDGE TYPE ______

MODEL: ______ POLYMER: ______ ______

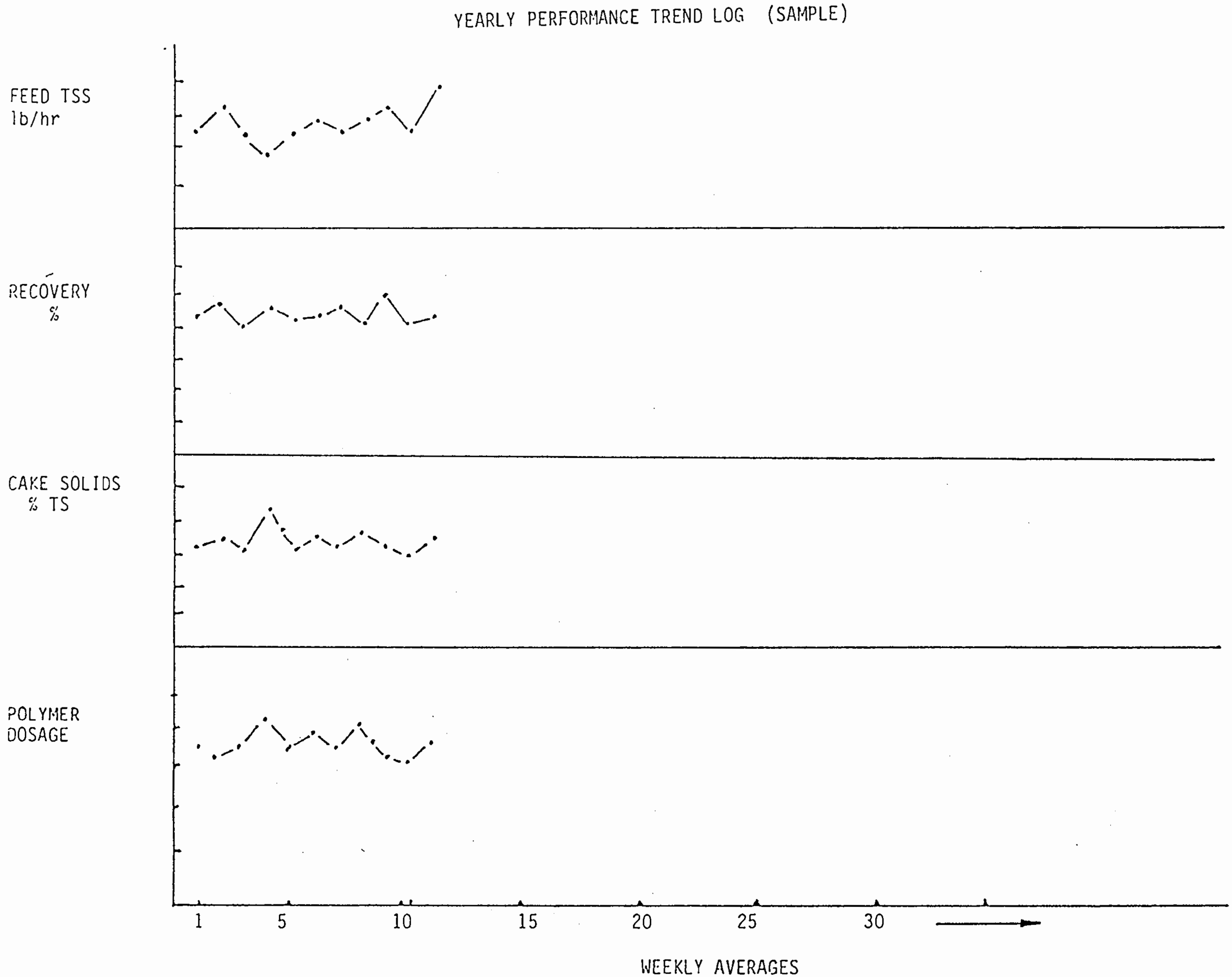


Figure B-4. Sample yearly performance trend log.

the aging and condition of the sludge prior to conditioning and dewatering is not possible. While ferric chloride and lime probably remain the primary choice in this situation, polymers are also being used quite effectively and at usually a reduced chemical cost. When using polymers, it is essential that the amount and kind of polymer required be evaluated on a routine basis.

The pour test, described elsewhere in this document, is particularly useful in providing a quick and quite effective evaluation of the polymer dose requirement. Generally one test early in the day is sufficient for those facilities where the sludge feed is relatively constant. However, in other cases more tests are needed to ensure adequate conditioning. Generally, mixed primary and waste activated sludge, or pure waste activated sludge without other treatment or conditioning history, will require a cationic polymer in varying doses. Most often the doses are in the range of 2.2 to 7.7 kg of active polymer/Mg of solids processed (2-7 lb/ton). In many cases polymer conditioning will not work unless the polymer is added directly to the feed sludge entering the press. That is, if not paced with the varying sludge feed, polymer conditioning will not work.

Since operating conditions (i.e., the SRT, F/M for the influent wastes, holding time, temperature, etc.) vary and may cause conditioning changes, it is necessary for the operator to keep a log of the polymer dosage required to reach a given degree of cake solids with a particular blend or a particular kind of sludge. If the dosage requirements change dramatically, it is the operator's responsibility to determine the origin of these differences. Possible explanations for these differences are listed below.

- When other sludge conditioning is practiced in the flowsheet, for example conditioning prior to flotation thickening, it may alter the effectiveness of downstream polymer additions, possibly to such a degree that additional treatment is required. In this case, often two polymers are required downstream from the first conditioning step to eliminate the residual effects of the first polymer.

- When lime, alum, or iron salts are used in the wet-end of the process (e.g., alum or iron salts is added in the primary tanks for phosphorus removal and the primary sludge is separately thickened and separately pumped to the dewatering operation), the makeup or blend of primary and waste activated sludge will change with time. Also, the quantity of alum or iron will vary in the same fashion. Sludges that have been treated with large quantities of either iron or aluminum in the wet-end of the treatment plant may require two polymers or, at a minimum, an anionic polymer for effective conditioning. Often the presence of these carryover materials will simply alter the quantity of cationic polymer required. Such wet-end conditions demand a routine, careful appraisal of polymer needs.

- The criteria that should be used to determine the correctness of the conditioning chemicals, whether inorganic or organic, are: 1) cake solids, 2) solids recovery, and 3) the rather subjective property of cake release from the cloth. Good cake release simply implies that the cake falls away cleanly from the filtration medium and does not penetrate into the medium or foul the medium so that continuous washing is required. Secondary considerations, such as how fast the filtration rate through a given cloth deteriorates, may also assume primary importance, particularly with iron and lime. If overconditioning is practiced on a long- term basis, especially with lime, an excessive rate of cloth blinding can be expected. The cake solids objective will be a function of the conditioning method and the blend of primary or waste activated sludge that is being dewatered. Recovery is a function of the same variables, but generally is not a significant factor unless cloths are torn.

  If the sludge being dewatered contains a high fraction of primary sludge (greater than 40 percent), the cake will discharge cleanly even with marginal or less than optimum conditioning practices. This is particularly true when using iron and iron salts and lime, but is also true when using polymer conditioned sludge. The greatest difficulty with cake release or cake discharge is usually encountered in operations dewatering 100 percent waste activated sludge, which is conditioned with polymer but without precoat or body feed.

Where the cake tends to be overly sticky or wet, the following factors should be considered prior to additional dewatering operations.

- Determine the optimum dose of the polymer being employed, and determine if the needs of the conditioning system are being met. For example, is polymer dosed at the actual feed rate?

- If body feed or precoat is used, check the timing of the application, quantity, and sequence of operation. Is there enough feed volume to apply the precoat evenly? Consult the manufacturer for further input.

- Routinely check (particularly those plants that have a history of possibly requiring two polymers) to see if the condition of the sludge has changed sufficiently to merit a change in polymer addition, types of polymer used, or sequence of addition.

- Be sure to follow the manufacturer's recommendations for mixing and aging of the polymer.

- If a polymer change has been recently introduced, make sure the active fraction of the new polymer is the same as that of the previous polymer. For example, some emulsion polymers are approximately 50 percent active; others are as low as 20-25 percent active. Failing to take into account the activity of the material, by modifying the calculation for the quantity of material, could reduce the actual dose to approximately 50 percent of the desired value, thus causing conditioning failure. The unit cost in $/ton should be the basis for selecting a polymer.

B.4.2 Determining Mass Balances

It is possible to make a complete mass balance by determining all of the parameters that may impact the quality of the cake or filtrate. In Figure B-5, a material balance for a filter press test is shown. Often, it is not practical to try to measure all of the filtrate volume, particularly with full-sized units, because of the quantity and relative inaccessibility of the filtrate system. Indeed, total quantity involved may make it impractical to capture all of the filtrate material.

It is possible to make a reasonable material balance without knowing the filtrate volume if a good composite of samples has been taken during the dewatering cycle to determine the solids concentration in the filtrate. If the filtrate is clear enough, one may assume 100 precent recovery. Generally, if the recovery is in excess of 98 or 99 percent, the mass balance calculations employing 100 percent recovery will not be seriously or adversely affected. In this case, one assumes that all of the cake and precoat solids are discharged with the cake and the loss to the filtrate is zero.

The following equations allow the designer to determine the mass balance quantities without determining the actual quantity or volume of filtrate.

$$\text{Feed solids rate} = 5\ Q_F\ F$$

where,

Q_F = feed rate, gal/min
F = feed solids, % TSS

$$R = 100\ [(C_S/F)]\ [(F-C_C)/(C_S-C_C)]$$

where,

R = recovery, % TSS
C_S = cake solids, % TSS (or TS)
F = feed solids, % TSS
C_C = centrate or filtrate solids, % TSS

B.4.3 Startup

At startup, sufficient jar tests should have been done to determine the optimum or near optimum chemical conditioning mode and the chemical requirements. It is essential that the feed of the required chemicals be the same as used in the calculations and the mixing be adequate to assure proper distribution of the polymer or other conditioning chemicals. It is much easier to achieve good mixing in a 1.0-l beaker than in a line carrying 12.6-126.2 l/s (200-2,000 gpm) of 4 percent sludge. If two chemicals are employed, such as ferric chloride and lime, or ferric chloride and a polymer, the ferric chloride should be applied first and at least 10 seconds of mixing time should be permitted prior to the addition of the second conditioning chemical. During startup, particular attention should be paid to: 1) flocculation and water release, 2) recovery, 3) cake solids, and 4) cake release, the primary objectives of the conditioning.

Cake release is one of those properties that is difficult to measure in a bench-scale test. It is possible in bench-scale work to determine whether cake release is quick, easy, and clean, or difficult. However, in the ranges between difficult and easy, it is hard to make anything other than a subjective kind of judgment. As the operator gains experience, his judgment will become more accurate. Usually cake release optimization must await full-scale operation, although broad range estimates of chemical requirements can be made on the basis of bench-scale tests.

Poor cake release results from the development of inadequate resistance to shear in the cake and, thus penetration of the cake into the pores of the filter medium. This usually occurs with a sloppy and inadequately dewatered cake. However, in some instances, it may occur with a cake that has been dewatered as well as possible, indicating the need to change conditioning procedures. If a sticky cake is encountered and chemical conditioning with ferric chloride and lime has been employed, the cake release may be enhanced or improved through the increase in the quantity of either iron, lime or both. It should be noted that iron is the coagulant and primary flocculant when this pair of chemicals is employed. Lime is used for flocculation, pH correction, and, at high lime doses, for structural stability.

When poor cake release is experienced with polymer conditioning, there are a variety of possible solutions that need to be examined, including:

- Alter polymer dose.
- Change the polymer to a similar but different type - that is, from a high molecular weight cationic to a medium molecular weight cationic.
- Examine the use of a multi-polymer system.
- Examine the use of ferric chloride and lime.
- Examine the use of body feeds or precoats.
- Extend the press time to reduce capacity.

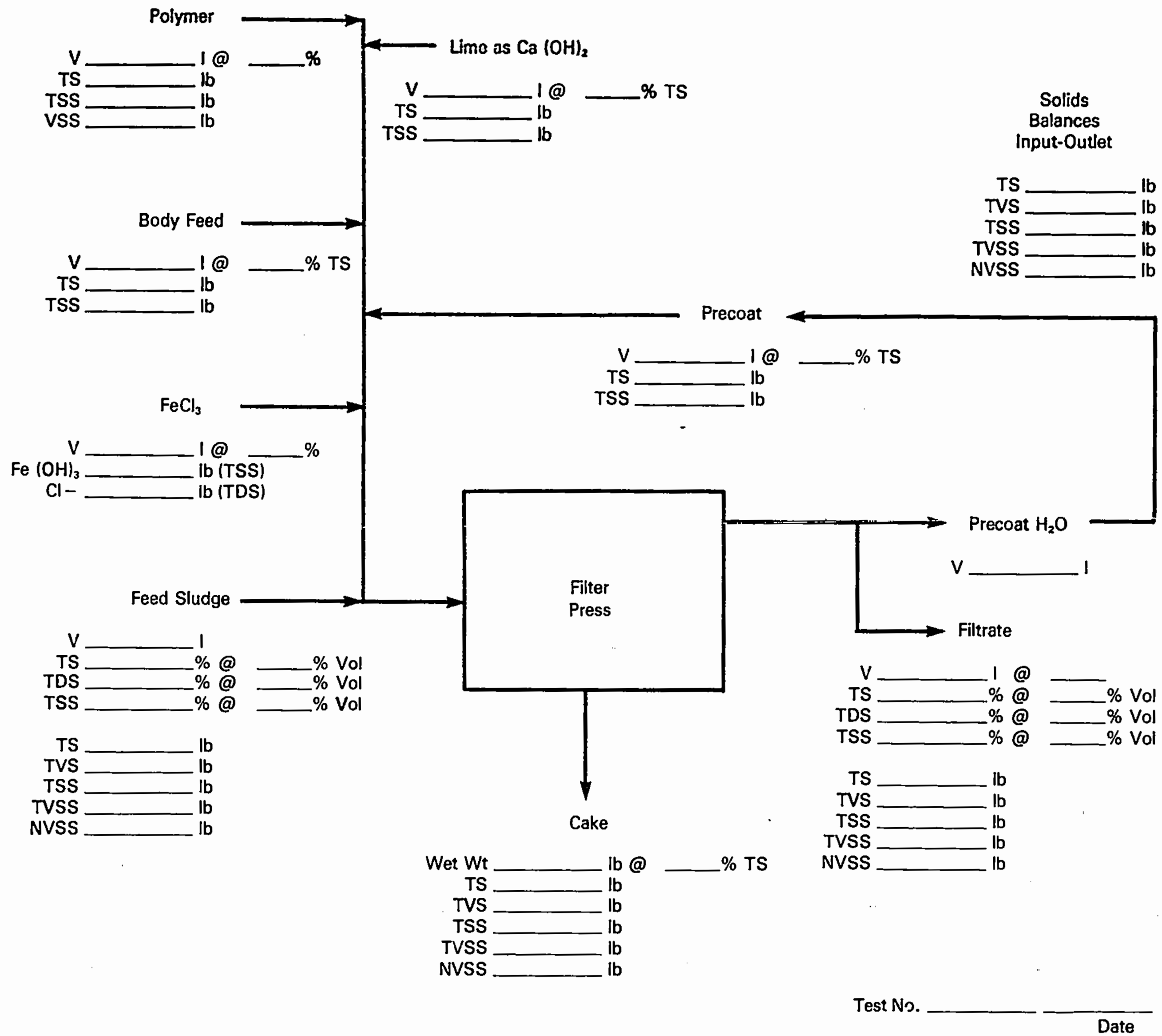


Figure B-5. Material balance - filter press test.

Actually, precoats and body feeds may be employed with ferric chloride and lime to increase their resistance to shear and to limit the penetration into the filtration medium. If a high volatiles solids content is desired, polymers should be employed because of the negligible alteration in the volatile fraction that they present. If sticking problems still persist after optimization of the polymer choice, it is likely precoat or a limited quantity of body feed or both may be the next best choice to maintain a high volatile fraction and improve cake release. If these fail, then it may be necessary to use inorganics to bring about the desired cake properties. p

As indicated earlier, record-keeping is absolutely essential for all operations where conditioning is required, and the rate of dewatering is an important consideration. A sample data sheet is provided in Figure B-6.

B.5 Vacuum Filters

This section will consider the operation of both drum filters and belt filters. These are the two common types of vacuum filters employed in the dewatering of domestic sludges. The differences between these two types of filters are outlined below.

- Drum filters have the cloth on the surface of the drum 360°.

- Drum filters usually employ doctor blades or a similar device to scrape the sludge from the surface of the filter medium.

- Belt filters have equipment for removing the filter medium from the surface of the drum immediately prior to discharge. Doctor blades, discharge rolls, or similar discharge devices can be employed. The removal of the belt from the filter drum also facilitates washing. The belt filters can be continuously washed in the area after cake discharge and prior to the cake form part of the cycle.

B.5.1 Sludge Conditioning System

For years, primary sludge and then primary plus trickling filter sludges were dewatered on vacuum filters using ferric chloride and lime. These operations were generally successful and produced an excellent cake that was manageable as a dry solid. However, since these operations used substantial concentrations of ferric chloride and lime, the mass of sludge was increased approximately 15 to 25 percent. Most new plants, existing plants, and plant modifications contemplate the use of polymer conditioning systems. Polymers can be successfully employed for the conditioning of 100 percent waste activated sludge on either drum- or belt-type filters. However, certain changes are required in the filter operation. Generally, operators try to achieve a final cake thickness in the range of 0.6-1.3 cm (1/4 to 1/2 in) when dewatering on a vacuum filter. This thickness is often associated with heat-treated sludges or sludges conditioned with ferric chloride and lime, or other agents for bulking in addition to lime.

There is no solids matrix to support the compressible waste activated sludge and to provide a channel for movement of water to the filter medium when conditioning with lime. Therefore, sludges high in biological material, typically 75-100 percent waste activated sludge, tend to collapse and lose their structure early in the filtering operation, leaving the outer portion of the cake extremely moist and resulting in a typical non-dischargeable filter cake.

To obtain a drier cake and a better cake release, several modifications in the filter operation must accompany the use of a polymer conditioning system:

- Generally, the cycle time must be shortened to effectively decrease the form time, hence cake thickness.

- If the cycle time cannot be shortened, the form time should be decreased by altering the extent of immersion of the vacuum filter in the sludge slurry. In this case, it may be necessary to change the bridging.

- It is almost always essential to increase the rate of vat agitation to avoid the problem of separation of solid and liquid phases in the vat. Further, it is often necessary to shear the larger flocs to restrict cake thickness.

These changes will generally result in a final cake thickness in the range of 0.3-0.6 cm (1/8 to 1/4 in). However, this cake will crack at the discharge roll of a belt filter and will normally be removed easily by a doctor blade on a drum filter. The cake is no longer too thick to collapse its own structure and it retains the water in the surface of the cake during the drying portion of the cycle. This moisture causes resistance to discharge by virtue of the physical features, that is, the sticky character of the cake and the penetration into the filter medium.

B.5.2 Determining Mass Balances

Mass balances may be accomplished by using the same kind of mass balance sheet and information provided in Section B.4 on filter presses. The operator should develop a mass balance sheet specifically for the type of vacuum filter used. Figure B-7 provides an example of a mass balance sheet.

Operating data sheets should contain full information on conditioning techniques and operational history of both the wet-end and dewatering system. An example of a data sheet is shown in Figure B-8.

Figure B-6. Sample test data sheet for filter press.

PRESSURE FILTRATION DATA

TEST

Test No: ____________

Test Date: ____________

Slurry Solids, wt%: ____________

Chamber Area, ft^2 ________

Cake Thickness, mm

Final: ________

Initial: ________

Dry Solids Density, g/cc: ________

Liquid Density, g/cc: ________

Data: g-ml ________ kg-l ________

Cake Weight:	Total	Part	Tare
Wet:	_____	_____	_____
Dry:	_____	_____	_____

Line # of Constant Pressure: ________

Output: Metric ______ English ______

	Time minutes	Pressure psig	Volume		Time minutes	Pressure psig	Volume
1.	_____	_____	_____	2.	_____	_____	_____
3.	_____	_____	_____	4.	_____	_____	_____
5.	_____	_____	_____	6.	_____	_____	_____
7.	_____	_____	_____	8.	_____	_____	_____
9.	_____	_____	_____	10.	_____	_____	_____
11.	_____	_____	_____	12.	_____	_____	_____
13.	_____	_____	_____	14.	_____	_____	_____
15.	_____	_____	_____	16.	_____	_____	_____
17.	_____	_____	_____	18.	_____	_____	_____
19.	_____	_____	_____	20.	_____	_____	_____

Filtrate pH: ______ Filtrate Solids, mg/l: ______

CHEMICAL ADDITION:

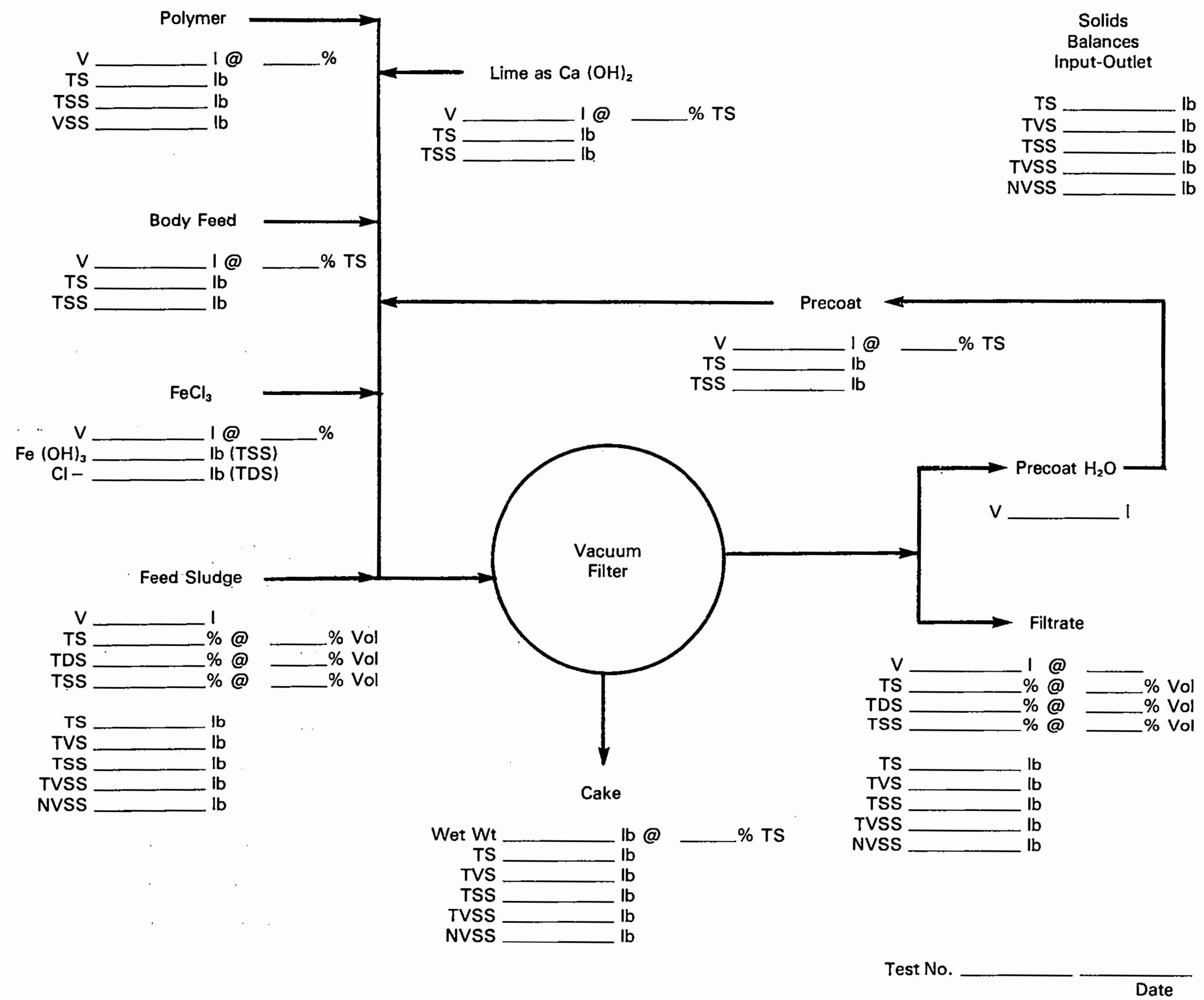
$FeCl_3$ ________

$CaOH_2$ ________

Polymer ________

Other ________

Figure B-7. Material balance - vacuum filter test.



Company ....

Address ....

....

TABLE NO. ....

Date Tested ....

By ....

Location ....

VACUUM FILTRATION TEST DATA SHEET

MATERIAL TO BE FILTERED:

.... % Suspended Solids, Consisting of ....

.... % Liquid, Consisting of ....

Slurry Temperature .... °C/°F.

Slurry pH ....

Specific Gravity of Liquid ....

% Dissolved Solids in Liquid ....

Filter Area .... Sq. Ft.

Filter Cloth ....

Slurry Feeding Technique: Bottom Feed ☐ Top Feed ☐

Air Flow Meter: Gas Meter ☐ Rotameter ☐

Avg. Filtrate Suspended Solids, ...Mg/L

TEST NO.	FILTERING VAC., IN. HG.			FILTERING TIME—MINUTES					AIR METER READING	FILTRATE VOLUME, ML.	WASH VOLUME, ML.	CAKE THICKNESS, INCHES	CAKE WEIGHT GRAMS		CAKE MOISTURE, %	AIR FLOW CFM/SQ. FT.	DRY WT. ×100 / WET WT. + FILTRATE WT.	W, DRY LBS. PER SQ. FT.				
	FORM	WASH	DRY	FORM	PRE-WASH DRY.	WASH	DRY	CAKE CRACKS					WET	DRY								
1																						
2																						
3																						
4																						
5																						
6																						
7																						
8																						
9																						
0																						

REMARKS: ____

Figure B-8. Sample test data sheet for vacuum filter.

The quantity of lime that will be in the solid rather than the liquid phase can be estimated by subtracting the estimated dissolved $Ca(OH)_2$ from the total amount added::

Dissolved $Ca(OH)_2$

$FeCl_3$ Neutralization, 70% of Fe dose

Dissolved phase from pH 7.0-10.0, 1.2 g/l

Solid phase - All remaining calcium will be present as ($Ca(OH)_2$ or $CaCO_3$ in the cake

B.6 References

1. Baskerville, R.C. et al. *Laboratory Techniques for Predicting and Evaluating the Performance of a Filterbelt Press*. Filtration and Separation 15(5):445, 1978.

2. *Design Information Report on Belt Filter Presses*. U.S. Environmental Protection Agency, Center for Environmental Research Information, Cincinnati, OH, 1985.

3. Belt Filter Press Survey Report. American Society of Civil Engineers, New York, NY, 1985.

4. *Standard Methods for the Examination of Water and Wastewater*. American Public Health Association, New York, NY, 1985.

Appendix C
Manufacturers and Sources of Equipment

C.1 Belt Filter Press Manufacturers

The list which follows is a compilation of many of the current (1987) belt filter press manufacturers. This list does not attempt to include all manufacturers of belt presses, consequently, some suppliers may not appear.

Arus Andritz Inc.
Arlington South Industrial Park
1010 Commercial Boulevard
South Arlington, TX 76017
817/465-5611

Ashbrook-Simon-Hartley Co.
11600 East Hardy
Houston, TX 77093
713/449-0322

Ralph B. Carter Co.
192 Atlantic Street
Hackensack, NJ 07602
201/342-3030

EIMCO
Process Equipment Co.
P.O. Box 300
Salt Lake City, UT 84110
801/526-2000

Envirex Inc.
A Rexnord Company
1901 South Prairie Avenue
Waukesha, WI 53186
414/547-0141

Komline-Sanderson Engineering Corp.
100 Holland Avenue
Peapack, NJ 07977
201/234-1000

Parkson Corp.
P.O. Box 408399
Fort Lauderdale, FL 33340
305/974-6610

Roediger Pittsburgh
R.J. Casey Industrial Park
Columbus & Preble Avenues
Pittsburgh, PA 15233
412/231-7979

C.2 Centrifuge Suppliers

The listing below includes many of the current suppliers of centrifuges in the United States. Since there are substantial differences in the bowl sizes and the gravitational forces for the various machines, Tables C-1 through C-4, inclusive provide this background information for three major manufacturers. The technical information provided in Section 7.3, used in conjunction with these tables, will assist in evaluating available and comparable machines. Capacity cannot generally be assigned to each machine since it is a function of the feed characteristics, product requirements, and other variables set forth in the technical discussions.

Alfa Laval, Inc.
2115 Linwood Avenue
Fort Lee, NJ 07024
201/592-7800

Bird Machine Company, Inc.
100 Neponset Street
South Walpole, MA 02071
615/668-0400

Broadbent Inc.
2684 Gravel Drive
P.O. Box 185249
Fort Worth, TX 76118
817/595-2411

Centrico, Inc. (Westfalia)
100 Fairway Court
Northvale, NJ 07647
201/767-3900

Clinton Centrifuge
P.O. Box 217, Dept. B
Hatboro, PA 19040
215/674-2424

Dorr Oliver, Inc.
77 Havemeyer Lane
Stamford, CT 06904
203/358-3200

GCI, Inc.
P.O. Box 217
220 Jacksonville Road
Hatboro, PA 19040
215/443-7878

Humboldt-Wedag
3260 Pointe Parkway
Atlanta, GA 30092
404/448-4748

Ingersoll-Rand, Inc. (Kruger)
150 Burke Street
Nashua, NH 03061
603/882-2711

Pennwalt Corporation
Sharples-Stokes Division
955 Mearns Road
Warminster, PA 18974
215/443-4000

Table C-1. Manufacturer A - Countercurrent Decanter Centrifuges

Bowl Size, mm	Maximum Gravities
460 x 1,370	3,500
610 x 1,830	3,000
610 x 2,440	3,000
760 x 2,440	3,000
760 x 3,050	2,100
915 x 2,750	2,100
915 x 3,660	2,100
1,120 x 3,350	1,980
1,120 x 4,470	1,980

Table C-2. Manufacturer B - High-G Centrifuges, Concurrent Design

Bowl Size, mm	Maximum Bowl Speed, rpm	Maximum Gravities
152 x 353	6,000	3,060
184 x 129	6,000	3,700
356 x 787	4,000	3,180
356 x 1,257	4,000	3,180
425 x 1,257	3,250	2,510
508 x 1,270	3,300	3,090
508 x 1,930	3,300	3,090
610 x 1,930	2,850	2,770
635 x 1,651	3,000	3,190
635 x 2,286	2,700	2,590
737 x 2,336	2,600	2,780
889 x 3,302	2,400	2,860
1,016 x 3,556	2,000	2,270

Note: Normal operation on sludge is at less than maximum bowl speeds shown.

Table C-3. Manufacturer C - Low-G Centrifuges, Concurrent Design

Bowl Size, mm	Maximum Bowl Speed, rpm	Maximum Gravities
250 x 750	4,400	2,680
350 x 950	2,250	990
450 x 1,350	2,325	1,360
530 x 1,400	1,965	1,440
530 x 2,200	1,450	620
600 x 1,800	2,050	1,410
600 x 2,500	2,050	1,410
900 x 2,500	1,600	1,290
1,100 x 3,300	1,265	980
1,400 x 3,300	1,100	950
1,800 x 4,400	860	740

Note: While maximum speeds are shown, best operation may be at speeds considerably lower than those shown.

Table C-4. Manufacturer C - High-G Centrifuges, Concurrent Design

Bowl Size, mm	Maximum Bowl Speed, rpm	Maximum Gravities
600 x 1,800	3,000	3,020
600 x 2,500	2,700	2,445
750 x 2,500	2,500	2,620
900 x 2,500	2,100	2,630
1,100 x 3,300	1,600	1,570

Note: While maximum speeds are shown, best operation may be at speeds considerably lower than those shown.

C.3 Filter Press Suppliers

The following list includes the major manufacturers of filter presses in the United States. This list does not attempt to include all suppliers of filter presses, consequently, some manufacturers may not appear.

Ametek, Inc.
Valley Foundry and Machine Division
2510 S. East Avenue
Fresno, CA 93706
209/233-6135

Clow Corporation
Box 68
Florence, KY 41042
606/283-2121

Dorr Oliver, Inc.
77 Havemeyer Lane
Stamford, CT 06904
203/358-3200

Duriron Company, Inc.
Box 1145
Dayton, OH 45401
513/226-4000

Edwards and Jones
17 Leslie Court
Whippany, NJ 07981
201/428-2828

Eimco PED
Box 300
Salt Lake City, UT 84110
801/526-2000

Envirex, Inc.
1901 S. Prairie Avenue
Waukesha, WI 53186
414/547-0141

Ertel Engineering Company
P.O. Box 3245
Kingston, NY 12401
212/226-6023

Hoesch Industries
Box 461
Wharton, NJ 07885
201/361-4700

Ingersoll-Rand
150 Burke Street
Nashua, NH 03061
603/882-2711

JWI Inc.
2155 112th Avenue
Holland, MI 49423
616/399-9130

Koppers Environmental
1900 Koppers Bldg.
Pittsburgh, PA 15219
412/227-2000

Kubota, America Environmental
405 Lexington Avenue
New York, NY 10174
212/490-8050

Netzsch, Inc.
119 Pickering Way
Exton, PA 19341
215/363-8010

Passavant Corporation
P.O. Box 2503
Birmingham, AL
205/853-6290

Perrin William, Inc.
432 Monarch Avenue
Dept. B, Ajax
Ontario, CANADA L15 2G7
416/683-9400

R&B Filtration Systems
Division of Buderus Corp.
2211 Newmarket Parkway, Suite 150
Marietta, GA 30067
404/955-9335

Star Systems, Inc.
101 Kershaw Street
P.O. Box 518
Timmonsville, SC 29161
803/346-3101

Sperry, DR, Company
112 North Grant Street
North Aurora, IL 60542
312/892-4361

Treatment Technologies, Inc.
10 Poplar Road
Honey Brook, PA 19344
215/273-2977

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United States
Environmental Protection
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