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Dewatering of Mine Drainage Sludge

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February 1973

DEWATERING OF MINE DRAINAGE SLUDGE

Phase II

By

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ABSTRACT

This report is a study of various acid mine drainage sludge conditioning methods and dewatering systems and includes an acid mine drainage and sludge characterization program. During this characterization program four sludges were selected as being representative of the various types of sludges produced by the lime/limestone neutralization of acid mine drainage. Three of these sludges were produced by lime neutralization and one by limestone neutralization.

The conditioning methods studied were freezing, use of flocculants and use of filter aids. Freezing was studied as a method of reducing sludge volume. Flocculants and filter aids were studied as methods of improving filtration rates. Flocculants were also studied on a limited basis as an aid to clarification.

Six dewatering systems were evaluated:

1. Conventional Rotary Vacuum Filtration
2. Rotary Precoat Vacuum Filtration
3. Pressure Filtration
4. Porous Bed Filtration
5. Thermal Spray Drying
6. Centrifugation

No single dewatering system was found best for all acid mine drainage sludges. However, on the basis of cost, the most promising acid mine drainage sludge dewatering techniques appear to be centrifugation, conventional rotary vacuum filtration, and rotary precoat vacuum filtration.

This report is submitted in partial fulfillment of Grant 14010 FJX under the sponsorship of the Environmental Protection Agency and the Coal Research Bureau of West Virginia University.

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SECTION I

CONCLUSIONS

The major conclusions drawn from this study of four sludges produced by the lime/limestone treatment of acid mine drainage are:

1. The primary factors which affect the rate at which acid mine drainage sludge can be dewatered are the solids content of the sludge and the type of neutralizing agent used at the treatment plant.

2. Flocculants can improve the dewaterability of the acid mine drainage sludges tested by as much as 74 percent.

3. Filter aids can improve the dewaterability of the acid mine drainage sludges tested by as much as 11 percent.

4. It is technically feasible to dewater all sludges tested by all dewatering systems tested except as noted below.

5. Norton sludge, a limestone sludge, was not applicable to dewatering by rotary precoat vacuum filtration.

6. Conventional vacuum filtration was the least expensive method of dewatering Norton sludge and centrifugation was the least expensive method of dewatering Shannopin sludge. Centrifugation and rotary precoat vacuum filtration were the least expensive methods of dewatering Banning sludge.

7. Freezing acid mine drainage sludge and then thawing it significantly reduces the settled volume of the sludge.

8. Only systems utilizing a precoat or porous bed filtration produced a reasonably clear filtrate. All other systems would require recycling the effluent to the clarifier.

SECTION II

RECOMMENDATIONS

Additional research is needed to:

1. Optimize the concentration of flocculants used in dewatering systems in terms of costs rather than maximum filtration rates.

2. Generate better data and cost figures for the dewatering systems studied, especially conventional vacuum filtration and centrifugation, by pilot plant studies.

3. Optimize on a cost basis an entire treatment process including either a conventional vacuum filtration or centrifugation dewatering system.

4. Improve the efficiency of porous bed filtration by optimizing depth of bed flooding and studying methods such as the use of flocculants and filter aids to increase the rate of drying of the sludge.

5. Determine the optimum operating parameters and the costs of freezing acid mine drainage sludge.

SECTION III

INTRODUCTION

This is the final report on Environmental Protection Agency Grant 14010 FJX entitled "The Thickening and Dewatering of Precipitates from the Lime/Limestone Treatment of Mine Drainage." Work on this program was conducted at the Coal Research Bureau, School of Mines, West Virginia University.

One product of the neutralization of acid mine drainage by either lime or limestone is a precipitate (sludge) of less than 10 percent solids. Currently, this waste product is either permanently impounded in large earthen lagoons or pumped into mined out underground workings or abandoned surface mines.

In situations where underground disposal is not practical, the permanent impoundment of this precipitate is proving to be costly in view of rising land values. Also permanent impoundment requires the construction of large earthen lagoons which may be expensive to build, mar the landscape, create a potential hazard when abandoned and represent a very inefficient use of land.

The purpose of this project was to evaluate a number of methods of dewatering this precipitate in order to reduce the problem of its disposal. After studying methods used in the United States and abroad to dewater other sludges such as the sludge produced by pickle liquor neutralization and sewage sludge, a series of dewatering systems was selected. Each system was studied individually (generally by using bench scale equipment) as to the feasibility of its application in the dewatering of coal mine drainage sludge. All the systems were then compared on the basis of cost and overall efficiency.

SECTION IV

SLUDGE AND ACID WATER CHARACTERIZATION

Before a study of the dewatering and conditioning characteristics of various coal mine drainage sludges could begin it was necessary to select a few sludges which would be representative of the range of sludge types commonly produced. Therefore field trips were taken to a number of treatment plants utilizing either limestone or lime neutralizing agents.

It was found that two factors affected the type of sludge that was produced from the treatment process. The first factor was the type of acid water being treated. Four types of coal mine waters were of interest. The first was an acid discharge, characterized by a low pH, high acidity and high mineral content (the mineral content as used in this report refers to the minerals in solution). The second water represented mine drainages with a pH of about 5.0 with most of the iron in the ferrous state. The third type of water represented mine waters that were very close to neutral and contained iron primarily in the ferrous state. The fourth type of mine water had a low pH, but contained most of the iron in the ferric state.

The second factor that influenced sludge characteristics was the treatment method. Twelve treatment plants were studied. Ten of these plants treated their water with hydrated lime. The other two plants treated their water with quicklime and rock dust limestone respectively. Samples were obtained from four of these twelve treatment plants to provide a fair cross-section of acid waters and treatment methods. The following treatment plants and their sludges were chosen for the conditioning and dewatering studies.

1. Shannopin No. 1 Airshaft Treatment Plant, Jones and Laughlin Steel Corporation, located at Bobtown, Pennsylvania.

2. Banning No. 4 Treatment Plant, Republic Steel Corporation, located at West Newton, Pennsylvania.

3. Norton Treatment Plant, Environmental Protection Agency, located at Norton, West Virginia.

4. Edgell Treatment Plant, Consolidation Coal Company, located at Wyatt, West Virginia.

Shannopin No. 1 Air Shaft Treatment Plant

Shannopin Treatment Plant treated water from the Shannopin Mine which operated in the Pittsburgh bituminous coal seam using continuous mining methods. The acid water treated was of the type which had a low pH, high acidity and high mineral content.

The Shannopin Mine No. 1 Air Shaft Treatment Plant treated 700,000 gallons per day (gpd) of acid water and consumed 5 to 6 tons of quicklime per day. A schematic diagram of this treatment plant is presented in Figure 1.

Acid water was pumped from the mine into a 3,500,000 gallon holding pond that had a retention time of five days. Part of the acid water flowed by gravity to a slake tank where it was mixed with quicklime which was also fed by gravity using a screw feeder. The slaked lime was mixed with acid water in a small mixing tank creating a lime slurry that had a pH of between 11 and 12. The lime slurry flowed from the mixing tank into a sluiceway. Approximately 20 feet along the sluiceway, after leaving the mixing tank, two pipes added more mine water from the holding pond to the lime slurry lowering the pH to approximately 8.5. Since the sluiceway was about 200 yards long the lime slurry and mine water were extensively mixed and further aerated as they flowed toward the settling pond.

When the plant was first constructed, a surface aerator was used, but it was found that sufficient ferrous oxidation occurred in the holding pond to make further aeration unnecessary. After mixing, the slurry flowed into one of two large settling lagoons (30,000,000 gallon capacity and 72,000,000 gallon capacity). When one settling lagoon was being used, the other lagoon was drained of as much water as possible and the sludge allowed to dry and compact. The two lagoons were alternated in this manner and the sludge permanently impounded in these lagoons.

Banning No. 4 Treatment Plant

Banning Treatment Plant treated water from the Banning No. 4 Mine which worked the Pittsburgh seam using continuous mining methods. The acid water treated was fairly representative of the type with a pH of around 5.0 and most of the iron in the ferrous state. The pH of this water was, however, found to be lower than 5.0, averaging around 3.1. This plant operated on a four day per week, twenty-four hour per day schedule and treated 2,200 gpm of acid water using hydrated lime at the rate of one-half ton per hour. Figure 2 is a diagram of this treatment plant.

Banning started its operation by mixing treated water with hydrated lime to form a two percent lime slurry. Mine water was pumped directly from the mine sump to the aerator tank where it was combined with the lime slurry. A pH probe located in the aerator tank regulated the flow of

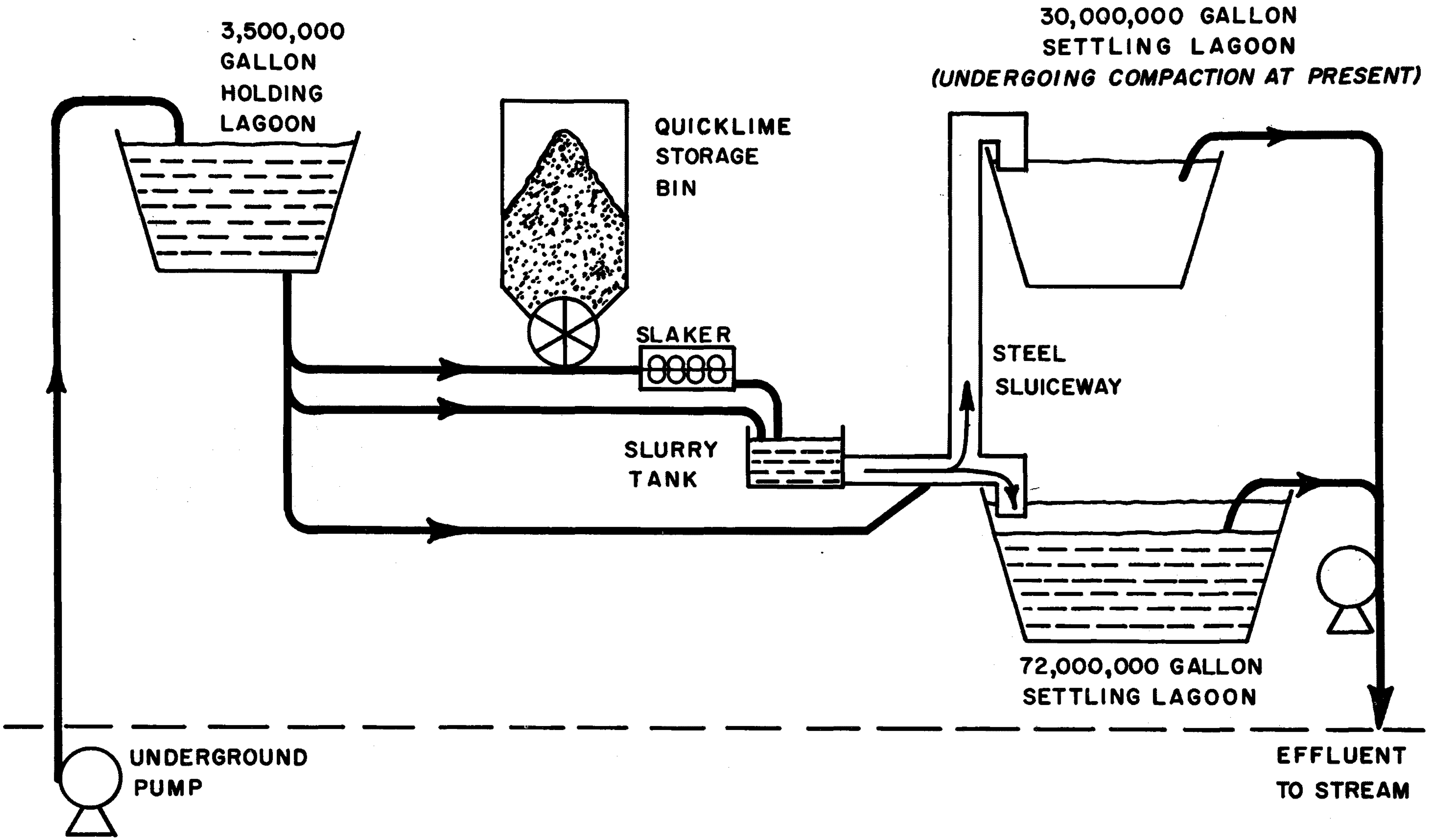


Figure I - SHANNOPIN TREATMENT PLANT SCHEMATIC DIAGRAM.

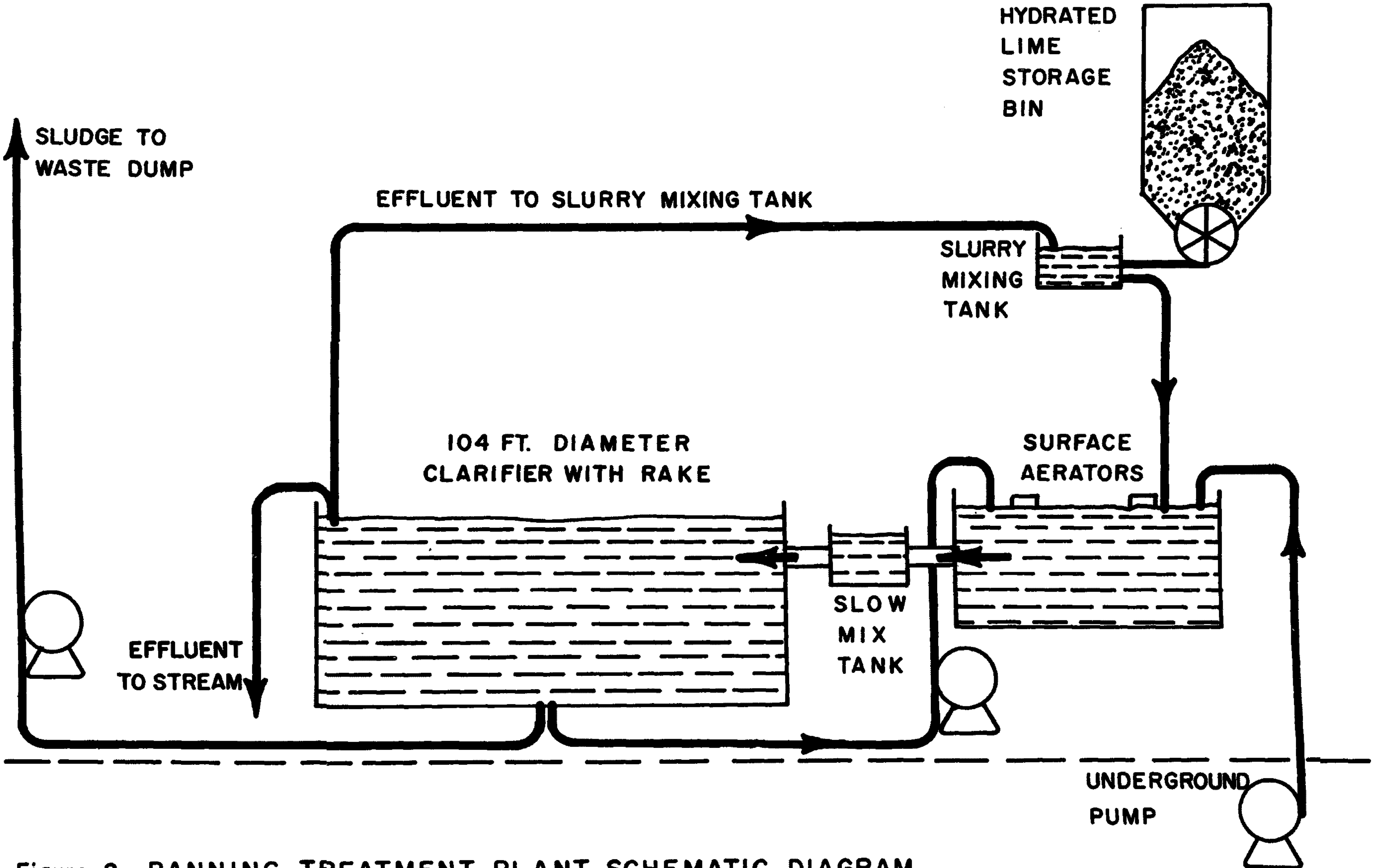


Figure 2 - BANNING TREATMENT PLANT SCHEMATIC DIAGRAM.

lime slurry to maintain a treatment pH of approximately 8.5. Two floating aerators mixed the acid water and lime slurry while aerating the mixture. After 30 minutes retention time in the aeration tank, the slurry flowed to a slow mix tank and then into a 104 foot diameter thickener clarifier. The settled sludge was moved to the center of the thickener by a rake mechanism. The sludge was then pumped for permanent storage into a settling basin that was also being used to store coal refuse. Part of the settled sludge (30-40 percent) was recirculated to the aeration tank.

Norton Treatment Plant

Acid water treated at Norton was pumped from the Grassy Run stream which was heavily polluted from abandoned coal mines. The water from this stream was highly acidic (pH 2.8) with nearly all of the iron in the water in the ferric state. This treatment plant, operated by the Environmental Protection Agency, was used for experimental purposes only. A diagram of the treatment plant is presented as Figure 3.

The acid water was pumped from Grassy Run through a sand filter into a 500 gallon holding tank. From the holding tank the acid water was then pumped at the rate of 15 gallons per minute into two 150 gallon mixing tanks where the water was mixed with limestone rock dust. The limestone addition was regulated by a pH recorder which maintains a pH range within the mix tank of 4.9 to 5.1. The treated water from each mix tank was then pumped into its individual 11,700 gallon settling tank. Total retention time for this system was approximately 14 hours. The treated effluent was then drawn off the settling tank and flowed back into the Grassy Run stream. Sludge was pumped off the bottom of the settling tanks into a 11,700 gallon sludge holding basin for further dewatering. Final sludge disposal was accomplished by dumping the sludge into mined out workings for perpetual storage.

Edgell Treatment Plant

Mine water at the Edgell Treatment Plant was pumped directly from a sump located at the western end of the Williams Mine. The Williams Mine was being operated by continuous mining methods and was mining part of the Pittsburgh coal seam. The acid water was of the type which was nearly neutral and contained iron primarily in the ferrous state. Part of the untreated mine water was used to mix the slurry in the slurry mix tank (lime consumption 480 pounds/hour) and the remainder of the mine water was directed into the flash mixer for complete neutralization (see Figure 4). The treatment pH for this operation was 8.0 to 8.4. The neutralized water then flowed by gravity into an earthen aerating lagoon that had a 15 hp surface aerator. The aerating lagoon had two outlets which allowed the sludge and treated water to flow by gravity into

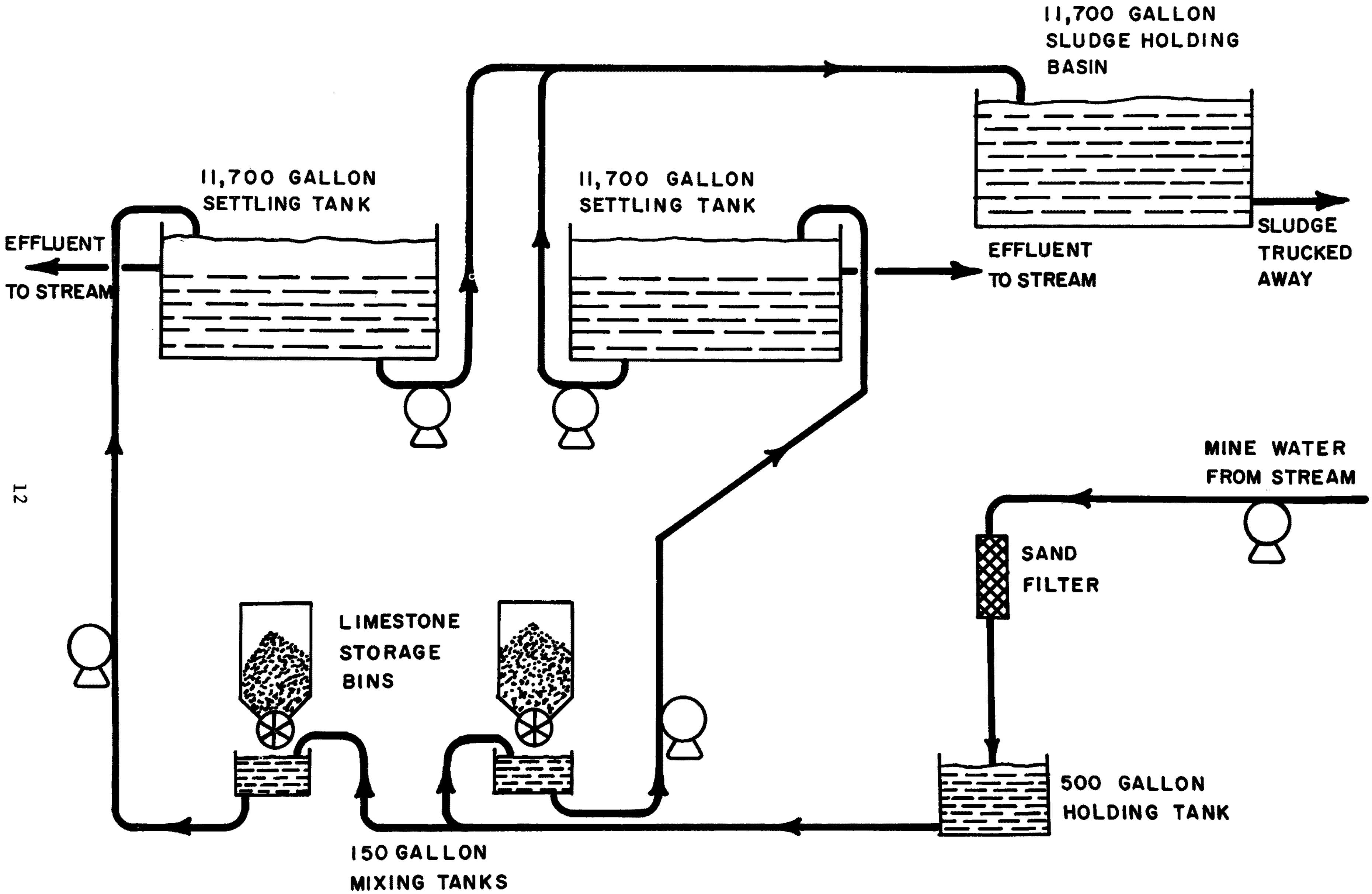


Figure 3 - NORTON TREATMENT PLANT SCHEMATIC DIAGRAM

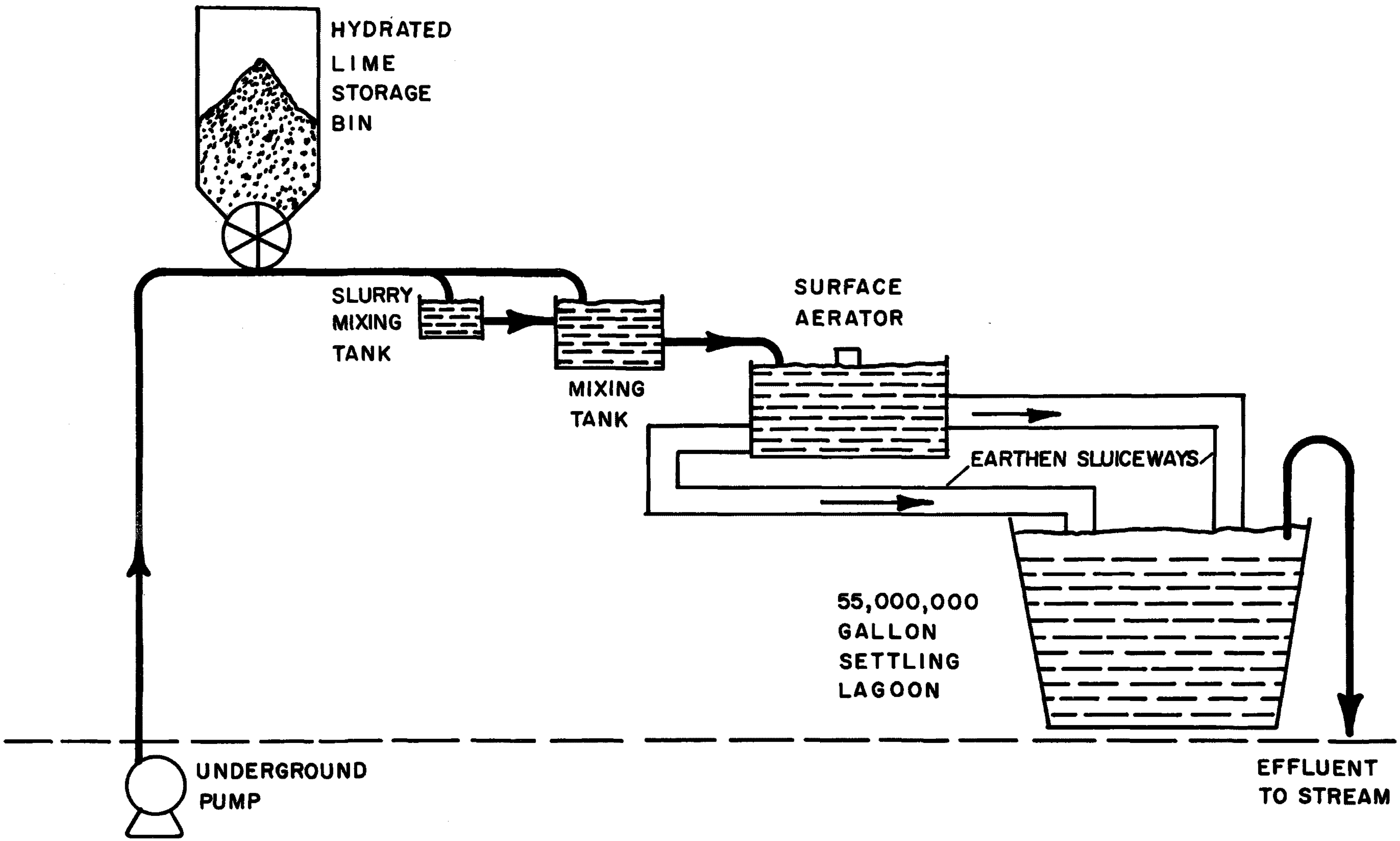


Figure 4 – EDGELL TREATMENT PLANT SCHEMATIC DIAGRAM.

separate ends of the settling pond. The settling pond had a capacity of 55,000,000 gallons and was used to perpetually store the sludge. The Edgell Plant treated mine water at the rate of 1,000 gallons per minute and ran on a twenty-four hour per day, seven day per week schedule.

Sampling Program

A sludge and acid water sampling and analysis program was conducted at each of the four characteristic treatment plants between June 22, 1970 and November 24, 1971. The purpose of this program was to provide information as to the variability of acid water and sludge at each treatment site and to collect sludge for the dewatering investigations. Since the primary purpose of this study was concerned with dewatering, no attempt was made to collect samples from the treatment sites at regular intervals, but rather samples were taken when additional sludge was needed for the dewatering studies.

Mine Water Sampling Procedure

Mine drainage chemical composition, especially the ferrous to ferric iron ratio, was very unstable and began to change upon formation. The main difficulty encountered in the sampling of mine drainage was keeping the iron constituents ratio stable long enough for the sample to be transported to the laboratory for analysis.

In order to completely determine the concentrations of the ferrous iron present in mine drainage, the following sampling and analytical procedure was devised. A mine water sample was taken at the point of treatment and was placed in a polyethylene bottle. The bottle was filled to its top to exclude as much air as possible. A second mine water sample was filtered through Whatman No. 40 filter paper into a polyethylene bottle that contained 100 milliliters (mls) of 3 normal hydrochloric acid. A third sample of mine water was placed into a polyethylene bottle that contained 20 mls of an acid mix of equal parts of sulfuric and phosphoric acid. All three mine water samples were returned to the laboratory for immediate analysis.

Mine Water Chemical Analysis Procedure

Mine water was analyzed for total iron, ferric iron, ferrous iron, silicon, aluminum, magnesium, calcium, acidity, alkalinity and sulfates in order to determine the major chemical parameters of the water. The chemical analysis was initiated in the field by determing the pH using a portable battery operated pH meter. The remaining analyses were performed in the laboratory.

The first water sample was immediately analyzed for acidity and alkalinity. Acidity analyses were performed by the Salotto Method.[1]

Alkalinity determinations were made using the method subscribed to by the American Water Works Association.(2) This latter method of analysis was used to determine the carbonate, bicarbonate and hydroxide alkalinity properties, and enabled the calculation of total alkalinity based on these results.

The second mine water sample had been filtered to remove all the precipitated ferric hydroxide in the raw water. The 3N hydrochloric acid added to the sample minimized further iron oxidation and precipitation. The sample was analyzed for silicon, aluminum, magnesium, calcium and total iron using a Perkin-Elmer Model 303 Atomic Absorption Spectrometer equipped with a DCR-1 digital readout device. Sulfates were analyzed using the barium chloride gravimetric method.

The third mine water sample which had been acidified with the sulfuric-phosphoric acid mixture was analyzed for ferrous iron by the potassium dichromate method described by Hall.(3)

The ferric iron was determined by subtracting the ferrous iron result from total iron as determined by atomic absorption.

Mine Water Chemical Analysis Results

Shannopin acid water was the most highly mineralized of the four waters studied as shown in Tables 1 through 4. Norton acid water was the lowest in mineral content. The Shannopin acid water was particularly high in aluminum and total iron relative to the other three waters. The Shannopin water, however, was relatively low in ferrous iron content presumably due to aeration or bacterial oxidation occurring within the holding pond. The Norton acid water was by far the lowest in ferrous iron as it was collected from a stream and was well aerated by natural processes.

No carbonate or hydroxide alkalinity was found in any of the samples; however, bicarbonate was detected in some Edgell samples.

The four acid waters studied represented a wide range of chemical composition and should therefore be reasonably representative of the range of mine drainage normally encountered. Mine water types of such quality as to not require treatment were not included in this survey

Sludge Chemical and Physical Analyses

The following information was determined in order to define the major physical and chemical parameters of the sludges.

1. Settling rate of slurries.
2. Percent solids of slurries and settled sludge.

Table 1

Shannopin Treatment Plant Chemical Analysis

Analysis of Raw Water

	High	Low	Mean
pH	2.9	2.7	2.8*
Ca (ppm)	430	400	410
Mg (ppm)	200	150	170
Al (ppm)	150	120	140
Si (ppm)	40	30	35
$SO_4^{=}$ (ppm)	5,000	4,300	4,500
Fe^{++} (ppm)	180	10	100
Fe^{+++} (ppm)	500	430	480
Total Fe (ppm)	670	510	580
Acidity (ppm $CaCO_3$)	2,400	1,800	2,000
HCO_3^- (ppm $CaCO_3$)	0	0	0

Analysis of Slurry

	High	Low	Mean
Ca (ppm)	2,100	1,400	1,600
Mg (ppm)	200	180	190
Al (ppm)	150	130	140
Si (ppm)	60	0	40
$SO_4^{=}$ (ppm)	5,000	4,400	4,600
Total Fe (ppm)	640	470	550
Nonfilterable Solids (ppm)	39,300	3,800	14,300

* Median Value

Table 2

Banning Treatment Plant Chemical Analysis

Analysis of Raw Water

	High	Low	Mean
pH	3.3	2.8	3.1*
Ca (ppm)	490	420	450
Mg (ppm)	140	120	130
Al (ppm)	40	25	35
Si (ppm)	30	20	20
$SO_4^{=}$ (ppm)	2,900	2,400	2,700
Fe^{++} (ppm)	180	140	160
Fe^{+++} (ppm)	90	50	60
Total Fe (ppm)	260	200	220
Acidity (ppm $CaCO_3$)	850	530	680
HCO_3^- (ppm $CaCO_3$)	0	0	0

Analysis of Slurry

	High	Low	Mean
Ca (ppm)	950	770	860
Mg (ppm)	130	120	130
Al (ppm)	50	40	45
Si (ppm)	25	25	25
$SO_4^{=}$ (ppm)	2,800	2,700	2,700
Total Fe (ppm)	300	210	260
Nonfilterable Solids (ppm)	3,400	1,400	2,400

---

* Median Value

Table 3

Norton Treatment Plant Chemical Analysis

Analysis of Raw Water

	High	Low	Mean
pH	2.9	2.8	2.9*
Ca (ppm)	160	100	120
Mg (ppm)	35	25	30
Al (ppm)	40	20	30
Si (ppm)	20	10	10
$SO_4^{=}$ (ppm)	1,000	600	800
Fe^{++} (ppm)	3	1	2
Fe^{+++} (ppm)	120	55	90
Total Fe (ppm)	120	60	90
Acidity (ppm $CaCO_3$)	670	360	520
HCO_3^{-} (ppm $CaCO_3$)	0	0	0

Analysis of Slurry

	High	Low	Mean
Ca (ppm)	500	180	310
Mg (ppm)	35	25	30
Al (ppm)	35	20	25
Si (ppm)	15	7	10
$SO_4^{=}$ (ppm)	1,100	610	790
Total Fe (ppm)	120	50	90
Nonfilterable Solids (ppm)	250	45	140

* Median Value

Table 4

Edgell Treatment Plant Chemical Analysis

Analysis of Raw Water

	High	Low	Mean
pH	6.7	4.6	6.0*
Ca (ppm)	480	280	370
Mg (ppm)	150	75	95
Al (ppm)	50	2	20
Si (ppm)	25	5	15
$SO_4^{=}$ (ppm)	5,900	4,100	4,800
Fe^{++} ppm	870	210	460
Fe^{+++} ppm	80	0	30
Total Fe (ppm)	870	220	490
Acidity (ppm $CaCO_3$)	1,700	50	660
HCO_3^- (ppm $CaCO_3$)	440	0	110

Analysis of Slurry

	High	Low	Mean
Ca (ppm)	1,400	620	920
Mg (ppm)	80	70	75
Al (ppm)	75	8	35
Si (ppm)	20	9	15
$SO_4^{=}$ (ppm)	4,300	4,200	4,200
Total Fe (ppm)	340	220	260
Nonfilterable Solids (ppm)	3,100	760	2,200

* Median Value

3. Chemical composition of slurries.

Sludge Sampling Procedure

Two types of sludge were taken during the sludge sampling and analysis program. The first sample was the sludge slurry that was formed by the neutralization process. This sample reflected the characteristics of the treatment plant at the time of sampling. These samples were taken prior to discharge in the settling lagoon. Settled sludge was the second sample taken and it was used in the evaluation of the various dewatering apparatus examined. In all cases, except for the Banning Treatment Plant, settled sludge was pumped from the bottom of the settling lagoon near the slurry discharge point in order to get as fresh a sample as possible. Settled sludge samples from the Banning Treatment Plant were taken from a bleeder line off the sludge recirculation system located in the main treatment plant building. All sludge samples were brought back to the laboratory in five gallon plastic jugs or 55 gallon drums.

The sludge sampling points for each treatment plant are summarized in Table 5.

Sludge Settling Tests

Settling tests were conducted in the field using 1,000 ml volumetric flasks, 1,000 ml graduated cylinders, and a timer. A settling test was initiated by filling a volumetric flask with slurry and immediately transferring the slurry to the graduated cylinder. The initial height of the slurry and starting time were recorded. As the sludge settled in the graduated cylinder, periodic readings of the sludge interface were taken and the data plotted.

Settling tests performed in the field were conducted to determine the sludge settling rate for the first few hours. Additional samples were brought back to the laboratory in order to determine the final settled volume. The rate of settling was also recorded for the samples returned to the laboratory in order to determine the effect of previous settling on the settling rate.

Sludge Solids Content

Percent solids determinations were performed routinely on the slurry and the settled sludge. The methods of analyses used for determination of percent solids of slurry and settled sludge (nonfilterable) were essentially the same as the procedures described in the Federal Water Pollution Control Administration Manual.(4)

Table 5

Sludge Sample Point Locations

Type of Sludge	Banning Plant	Edgell Plant	Norton Plant	Shannopin Plant
Slurry	At slow mix tank	At discharge point from aerator	At discharge points into settling tank	At discharge point in lagoon
Settled sludge	From bleeder line located in treatment plant	Off shore of settling lagoon	From settling tank	Off shore of settling lagoon

Sludge Chemical Analyses

Chemical analyses were performed on the slurry sludge. The purpose of the chemical analyses was to observe the change in sludge chemical constituents as the raw water changed. Iron, calcium, magnesium, aluminum, silicon and sulfates were routinely determined for all the sludge samples. Atomic absorption analysis was used for the determination of metallic ions after dry precipitates present in the sample were dissolved with hydrochloric acid. Sulfates were determined by the barium chloride gravimetric method. Tables 1 through 4 summarize the chemical data for each slurry.

Sludge Physical Analysis Results

Sludge treated at the Edgell Treatment Plant settled faster than the other two lime sludges examined; the next fastest being Banning sludge followed by the Shannopin sludge. Norton sludge did not settle with a distinguishable interface and so its rate of settling was not evaluated. From the settling tests conducted it can be concluded that most of the settling for each sludge was complete in three hours or less. The settling rates for the three sludges which settled with a distinct interface are shown in Figures 5 through 7. These figures also show a comparison between the settling rate for each sludge upon collection and its settling rate approximately 24 hours later when the sample was resettled in the laboratory.

Each sludge occupied a different volume after settling and compaction were complete.The Norton sludge settled to the lowest settled volume (1.1 percent). The Edgell sludge settled to the next higher volume, (4.0 percent) followed by Banning (8.0 percent) and then Shannopin, (25.0 percent). All sludge volumes are given in terms of percent of the original slurry volume.

The general settling properties of the sludges studied are similar to other mine drainage sludges reported in the literature. The limestone sludge settles to a very small volume which agrees with previous work conducted by Wilmoth et al.[5]; however, since the Norton water was the lowest in mineral content of those studied, its relatively smaller sludge volume cannot be attributed only to the use of limestone but also to the small amount of sludge forming minerals in the water. In the report by Wilmoth et al. a direct comparison was made between lime and limestone sludge. Norton sludge (the limestone sludge) was found to occupy only two thirds of the volume of the lime sludge. Similarly the Shannopin sludge, which is created from a highly mineralized mine water neutralized with quicklime, had a large final settled volume. As would be expected, sludge from Shannopin and Banning settled faster in the laboratory than in the field since any surface change on the labor-

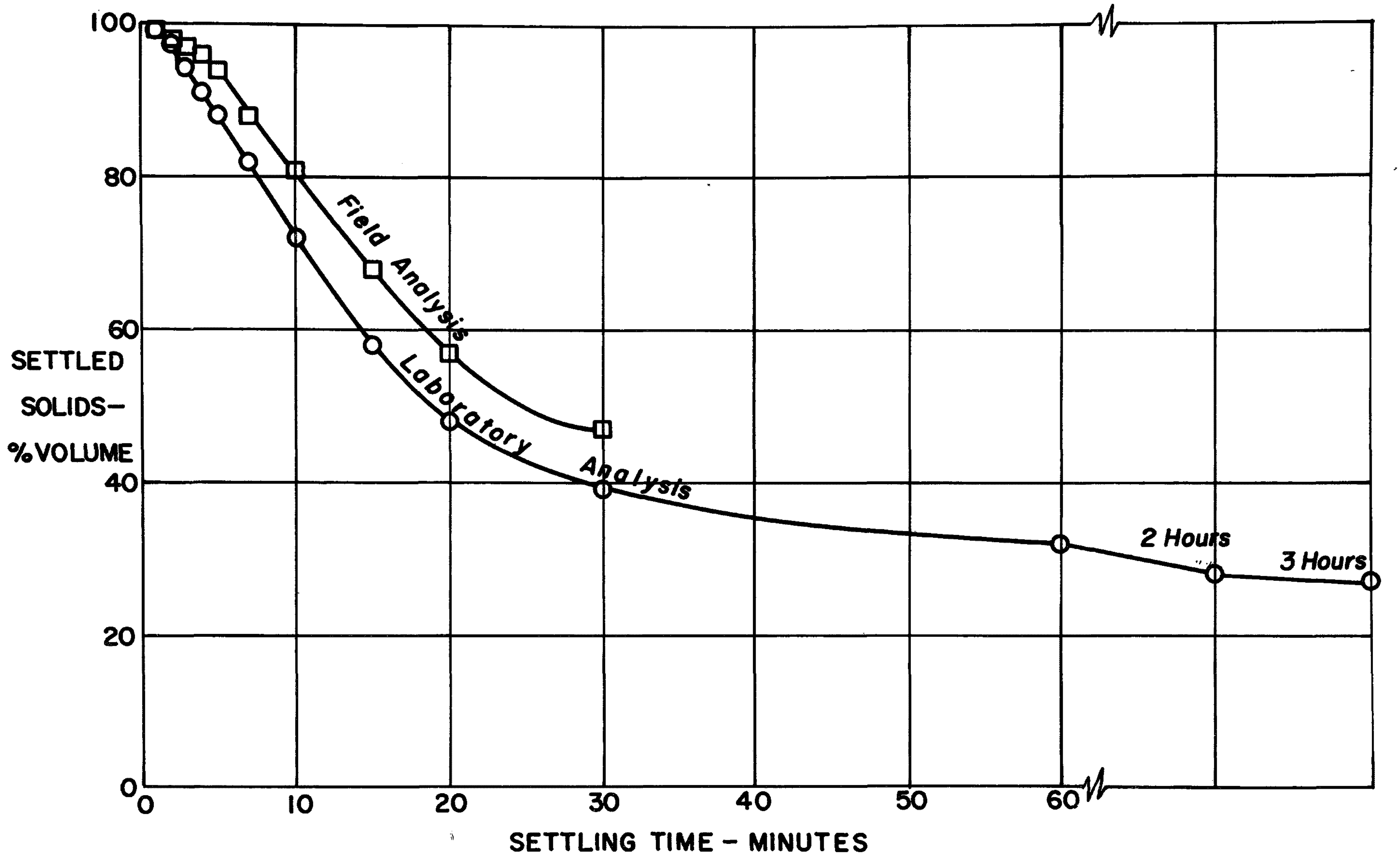


Figure 5- SETTLING RATE OF MIXING TANK SLUDGE FROM SHANNOPIN TREATMENT PLANT.

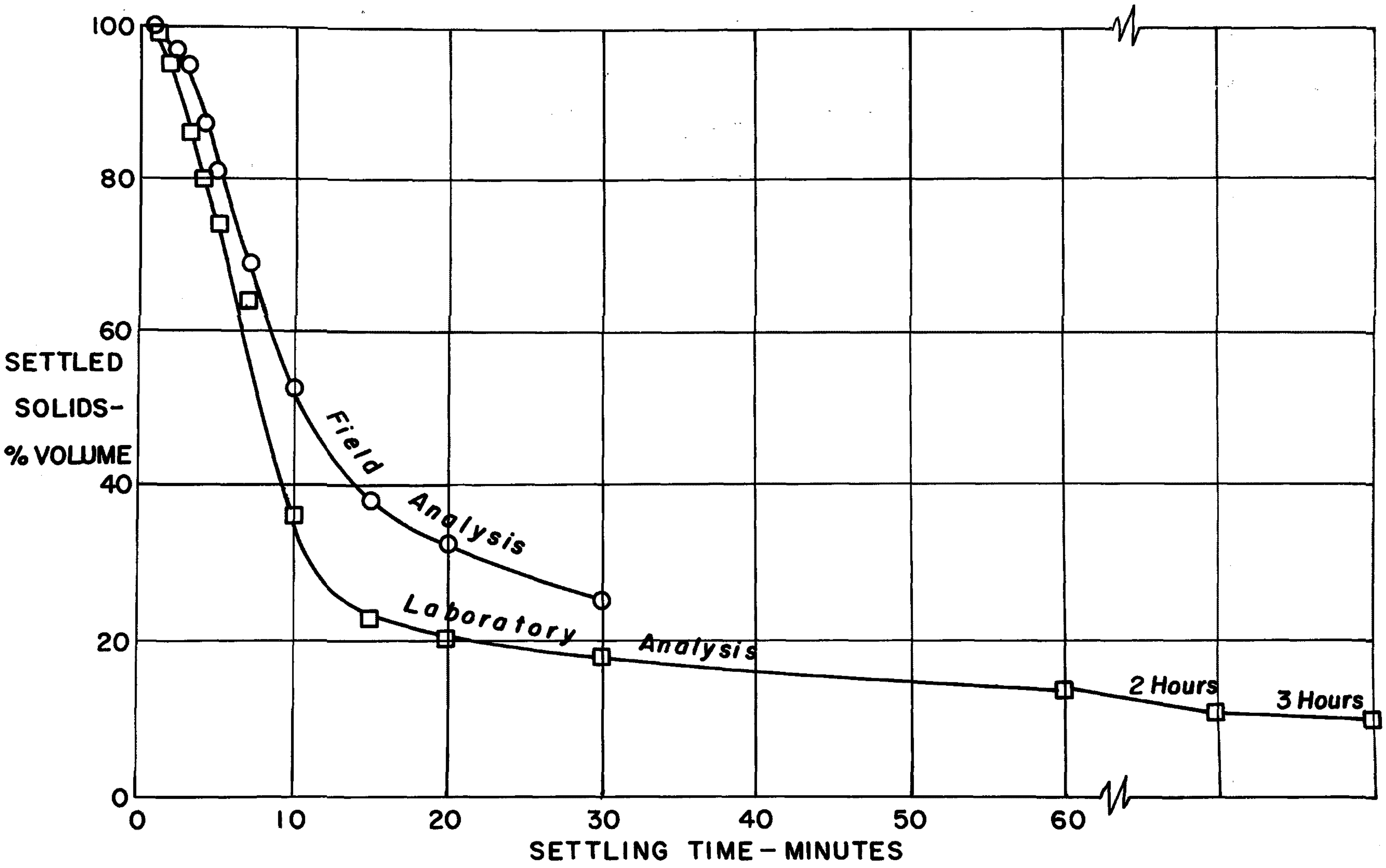


Figure 6- SETTLING RATE OF AERATOR SLUDGE FROM BANNING TREATMENT PLANT.

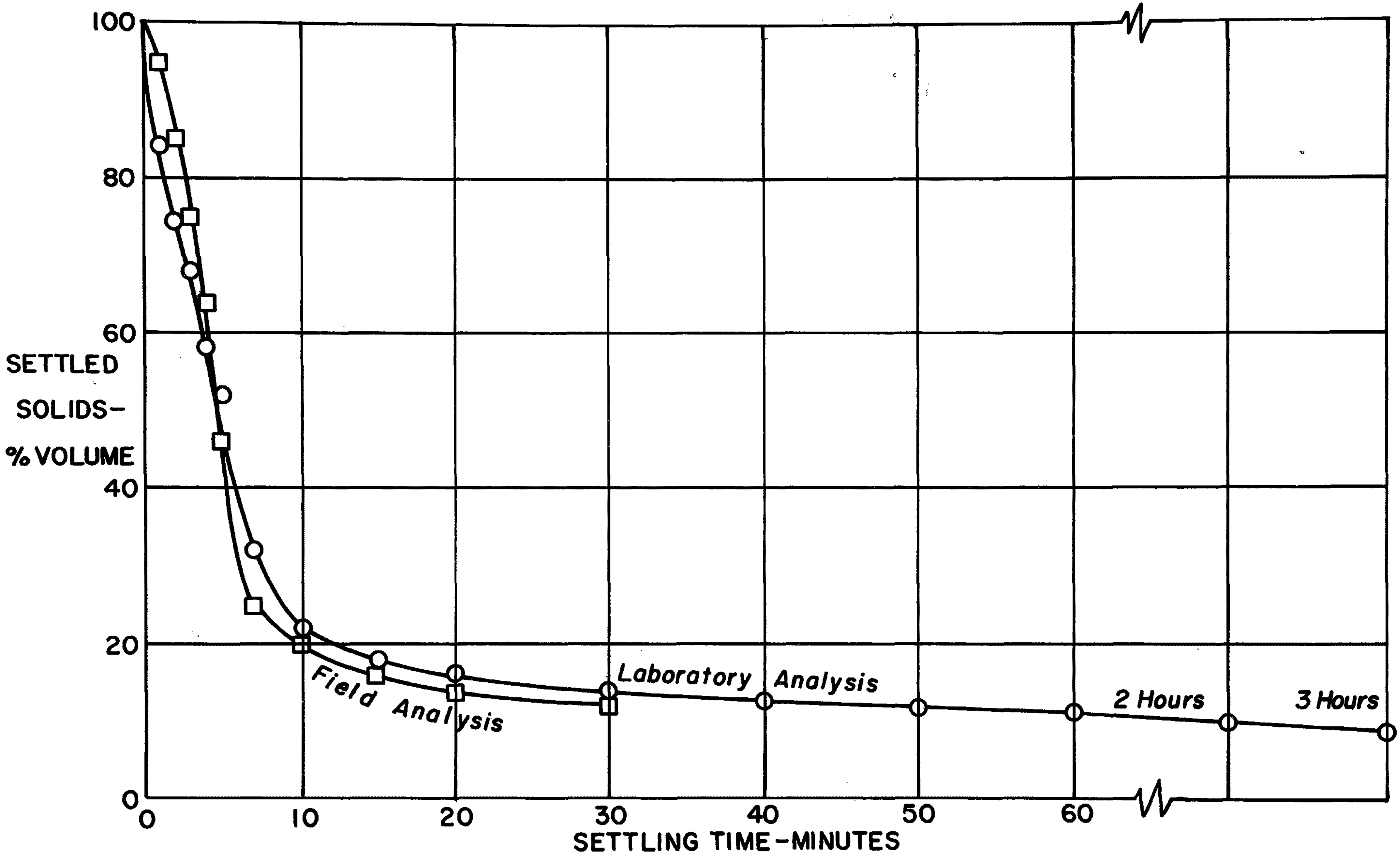


Figure 7- SETTLING RATE OF AERATOR SLUDGE FROM EDGELL TREATMENT PLANT.

atory sample had been dissipated by previous settling. Edgell sludge showed little change in settling rates after previous settling, indicating that settling had little effect on the physical character of the sludge particles.

Sludge Chemical Analysis Results

The results of the sludge characterization program are summarized in Tables 1 through 4. The Shannopin slurry was the highest in elemental concentrations of the four sludges studied and also was highest in nonfilterable solids. This was to be expected since Shannopin water had the highest mineral content of the four waters. The Norton slurry was the lowest in elemental concentrations and again this is to be expected since Norton water was the lowest of the four in mineral content.

The four slurries represented a fairly wide range of chemical composition and should therefore represent a fair cross section of the various types of sludges.

SECTION V

SLUDGE CONDITIONING

Several sludge conditioning methods were examined in order to reduce sludge volume and/or increase filtration rates during sludge dewatering processes. A unique conditioning process called sludge freezing was investigated as a method of reducing sludge volume and numerous flocculants and filter aids were investigated as conditioning agents to increase filtration rates. Since the purpose of this report was primarily to determine the feasibility and efficiency of various coal mine drainage sludge dewatering systems, flocculants were evaluated primarily as they related to dewatering. However, it was reasoned that if a flocculant was to be utilized at a treatment site for dewatering, it could be applied to the slurry before it entered the clarifier and simultaneously enhance both settling rates and dewaterability. Accordingly, clarification studies were also conducted on various sludges using flocculants which were selected with improved filtration rates as a criteria.

Sludge Freezing

General

Freezing, as a sludge conditioning process, has been investigated by a number of researchers with sewage and water works sludges. Early investigations found that, after freezing, sludge solids settled at a faster rate and settled to a smaller volume than did unfrozen sludge solids. The nature of the freezing process is not exactly known; however, freezing appears to destroy the gelatinous structure of the sludge allowing the entrapped water and solids to separate.

The early research into sludge freezing has resulted in the construction of several plants that freeze conditioned waterworks sludge. Due to the relatively high cost per unit volume of freezing, secondary sludge thickening was especially attractive and is generally utilized at these plants. The designers found, for example, in one plant that by slow stirring, the quantity of sludge to be frozen could be reduced from 33,000 gpd at .5 percent solids to 6900 gpd at 2.4 percent solids. Following the thickening process the sludge was pumped to a freezing tank where it was frozen and then thawed. The sludge solids and liquid were then allowed to separate by gravity draining.[6]

From the early research it was found that sludge must be completely frozen but at a relatively slow rate.[7] Recent studies on sewage sludge have shown that sludge freezing can be achieved by using the film-freezing principle. Film freezing of sludge operates on principles similar to extended freezing but freezing time is reduced since the sludge is frozen as a thin film.[8]

The cost of freezing waterworks sludge was found to be high. In one case, the freezing cost was $6.78 per 1000 gallons of sludge as compared to $5.04 per 1000 gallons of sludge dumped into lagoons. One of the major justifications of the extra cost was the use of the land for agricultural purposes that would have otherwise been used for lagoons.[9]

Test Equipment - Description

To observe the effects of freezing on different types of coal mine drainage sludge, laboratory scale freezing tests were conducted on the four characteristic sludges.

Equipment used for this series of experiments consisted of a conventional freezing compartment of a household refrigerator, plastic beakers, graduated cylinders, a thermometer, and a timing device.

Test Procedure

In order to observe the general effects of freezing on sludge properties, a standard was established for comparison. This was accomplished by taking 500 ml samples of the 4 sludges described in the sludge and acid water characterization section, allowing each to settle for six hours, and determining the final settled volume of the sludge. Following the establishment of a standard unfrozen settled volume for each sludge, 500 ml samples of the four sludges were then introduced into a freezing environment (-14°C) for 4, 5, 6, 7, 8, and 24 hours. After freezing, each sample was allowed to thaw and then was reintroduced into a 500 ml graduated cylinder. The sludge was allowed to settle for six hours and the final settled volume was determined along with percent solids of the settled sludge.

Test Results

Artificial freezing was found to reduce the volume of coal mine drainage sludge. Table 6 summarizes the results from the sludge freezing experiments.

Freezing appeared to have the greatest effect on the sludges that were produced from lime treatment. This was evidenced by the similarity in the reduction of settled sludge volume (approximately 90 percent) following freezing from the Shannopin, Banning and Edgell sludges. These plants used either hydrated lime or slaked quicklime for water treatment. However, in the case of the Norton sludge which was

Table 6

Summary of Results from Sludge Freezing Tests

Treatment Plant	Sludge Solids Content After Freezing (percent)*	Reduction in Settled Sludge Volume After Freezing (percent)**
Norton	21.0	47.0
Edgell	17.8	90.6
Banning	6.3	90.5
Shannopin	13.9	88.5

* 6 hours of settling and water over sludge removed

** 6 hours of settling

treated with limestone, a reduction in settled volume of only 47.0 percent was observed.

The reasons for the substantially greater decrease in settled sludge volume after freezing for lime sludges compared to the limestone sludge are not exactly known. However, it is known that the sludges created from lime treatment are substantially more gelatinous in structure and hold more water than sludges from limestone treatment. Since the effect of freezing is thought to be due to a breaking down of the gelatinous structure of the sludge, a greater degree of volume reduction following freezing would be expected with lime sludges.

Figures 8, 9, 10 and 11 illustrate the relationship of degree of freezing to settled sludge volume. It can be seen from the graphs that as the time the sludge was in the freezing environment increased, the settled volume of the sludge generally decreased.

The graphs, to a certain degree, suggest a near linear relationship between the degree of freezing and final settled sludge volume at least for the first portion of the curves.

The time necessary to completely freeze the quantities of sludge that were studied is undoubtedly less than 24 hours. It was the intention, however, to guarantee complete sludge freezing, therefore, an arbitrary 24 hour maximum freezing time period was chosen.

It can be concluded from the graphs that the full effects of freezing do not take place until the sludge is completely frozen. This fact agrees with research conducted on other types of sludge.[9]

Accompanying the substantial reduction in sludge settled volume following freezing was the requisite increase in sludge percent solids as shown in Table 6. The final percent solids of the frozen sludge was determined following six hours of settling with the clarified water over the top of the sludge removed.

To summarize, artificial freezing of coal mine drainage sludge can in most cases substantially reduce the final settled volume and concurrently increase the sludge solids content. The research conducted on freezing of coal mine drainage sludge suggests that the success achieved with a commercial waterworks sludge freezing process could be emulated with coal mine drainage sludge. The ultimate application will, of course, relate to the economics of the process.

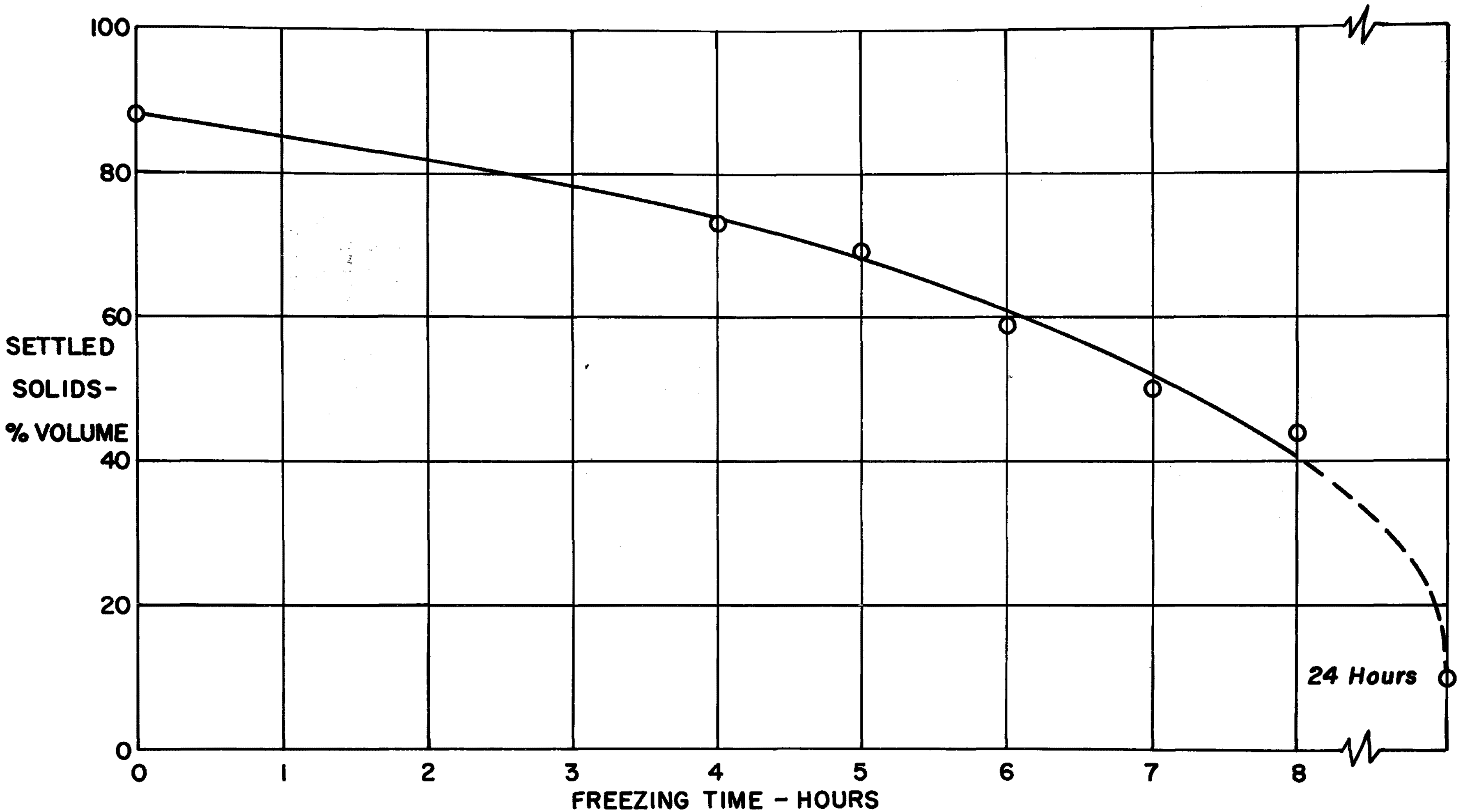


Figure 8- RELATIONSHIP OF FREEZING TIME AND VOLUME OF SOLIDS FOR SLUDGE FROM SHANNOPIN TREATMENT PLANT.

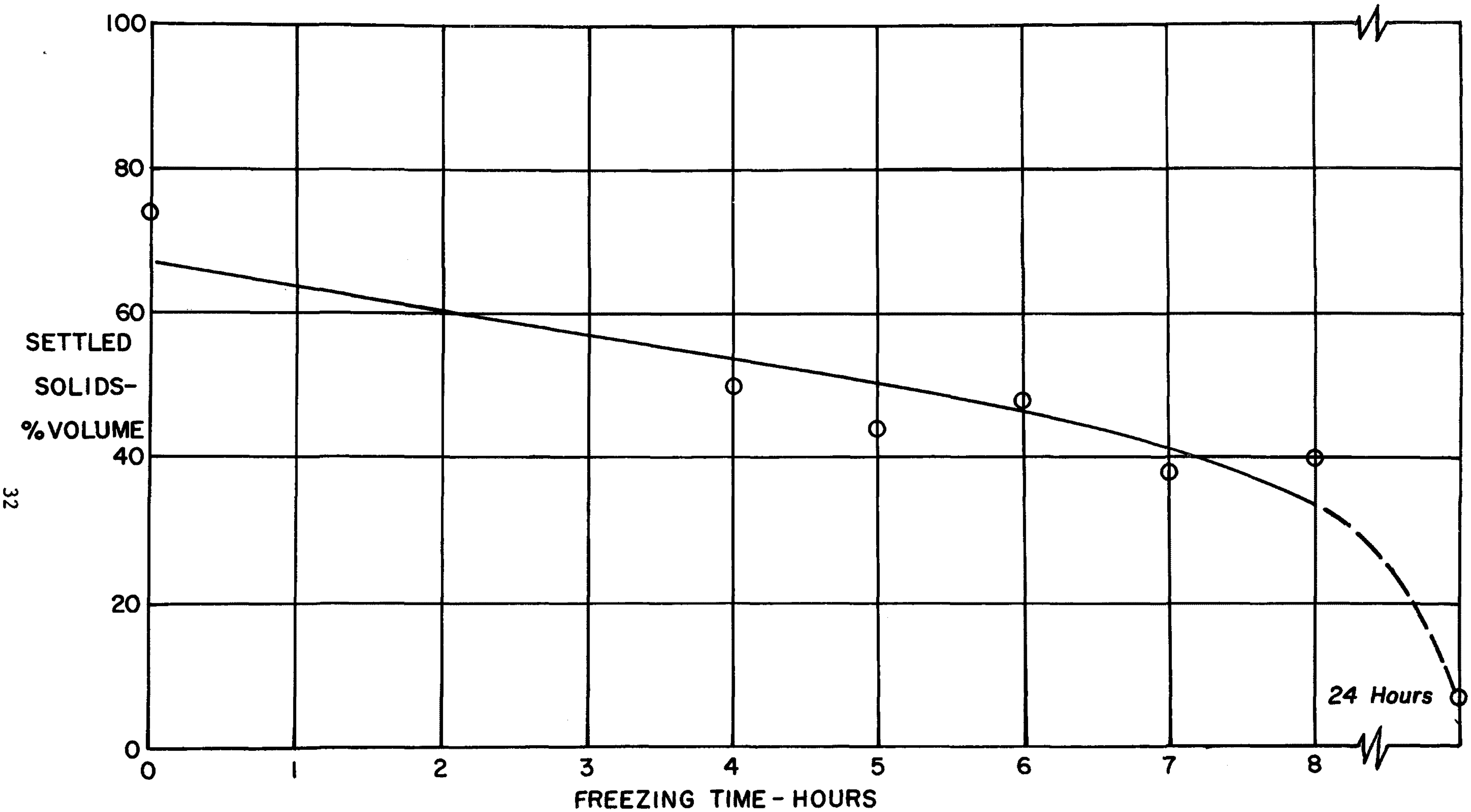


Figure 9 - RELATIONSHIP OF FREEZING TIME AND VOLUME OF SOLIDS FOR SLUDGE FROM BANNING TREATMENT PLANT.

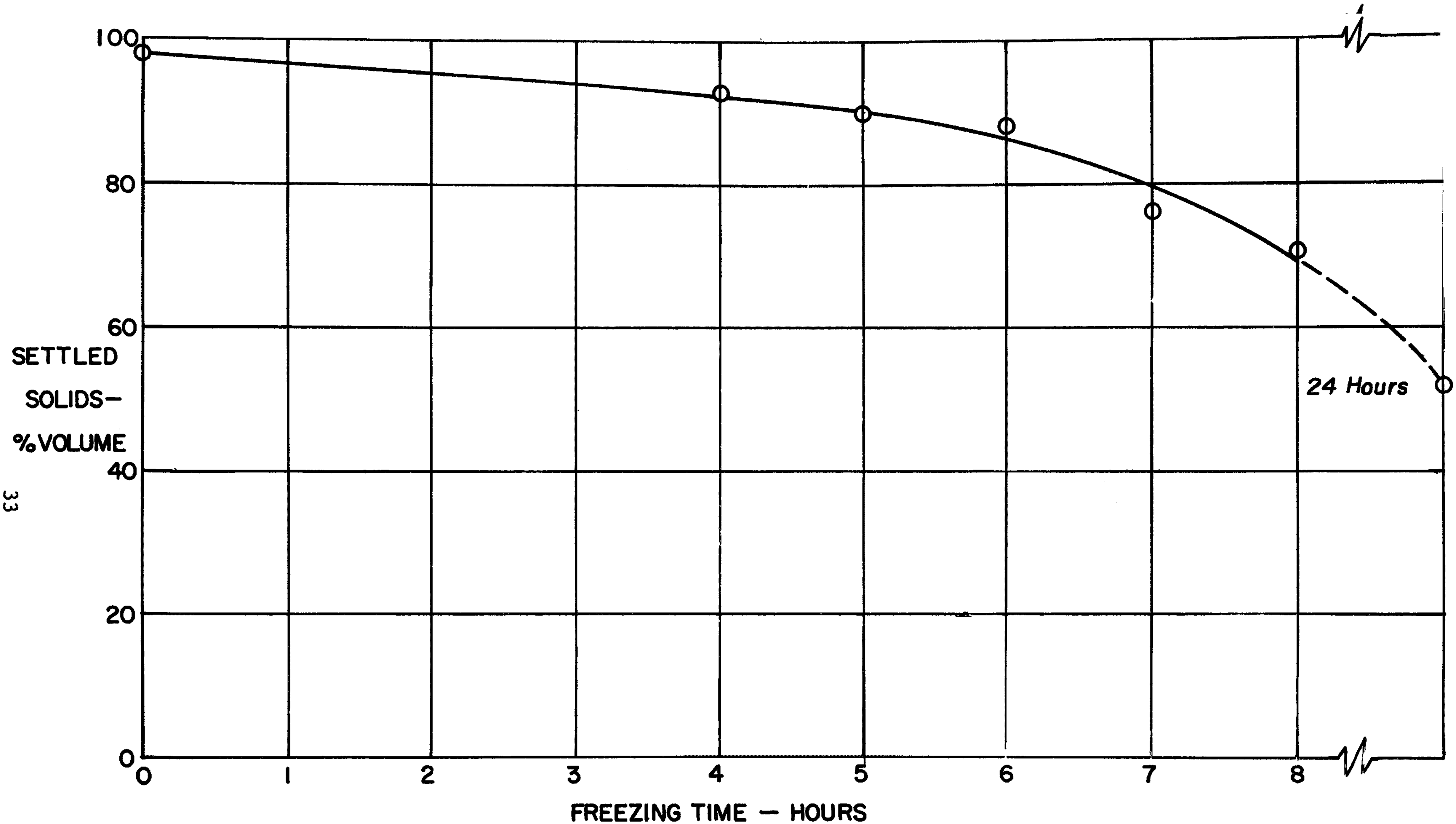


Figure 10— RELATIONSHIP OF FREEZING TIME AND VOLUME OF SOLIDS FOR SLUDGE FROM NORTON TREATMENT PLANT.

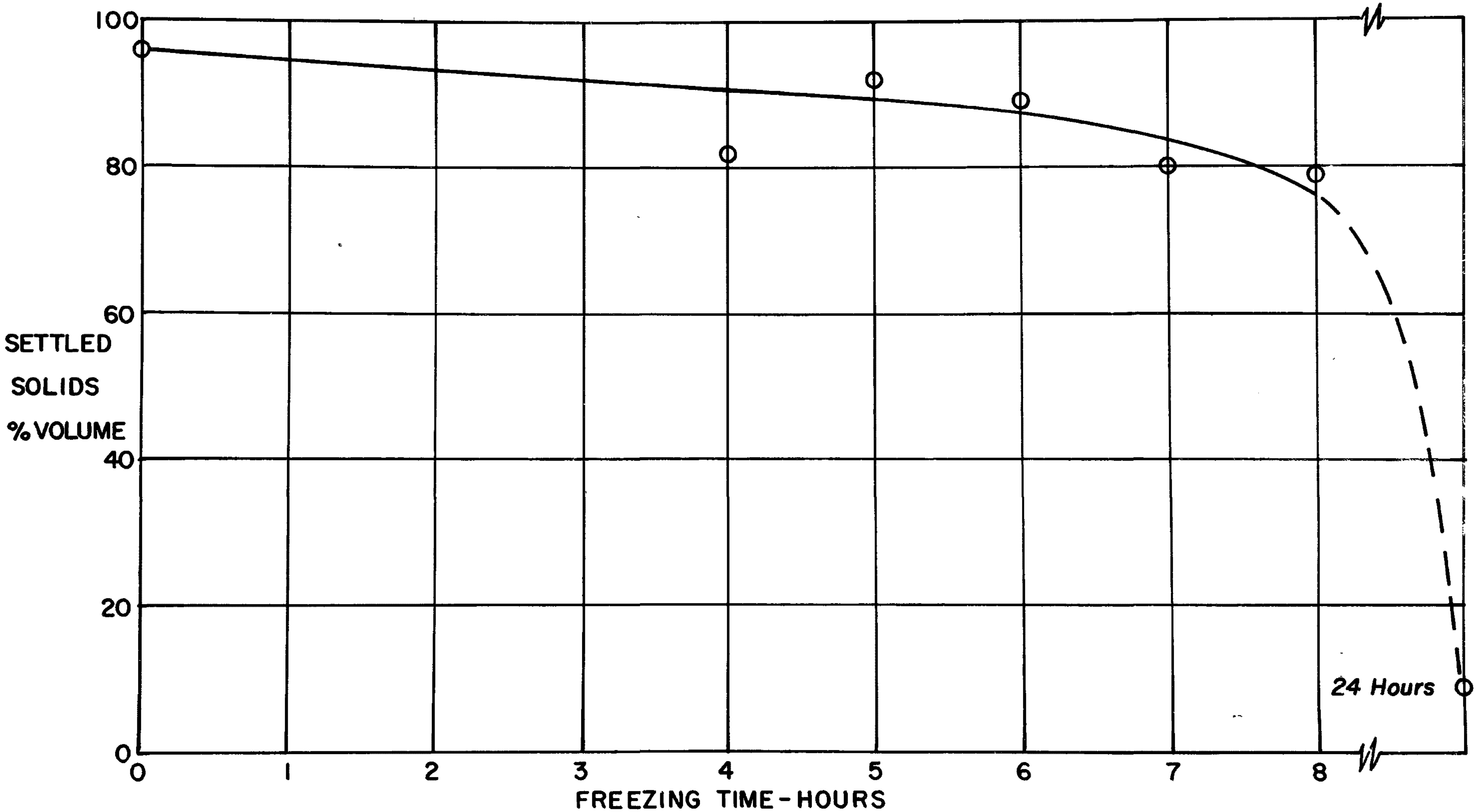


Figure 11 - RELATIONSHIP OF FREEZING TIME AND VOLUME OF SOLIDS FOR SLUDGE FROM EDGELL TREATMENT PLANT.

Conditioning For Dewatering and Clarification

General

Chemical conditioning (the use of chemicals to alter sludge characteristics) of sewage and industrial sludges has been studied by various workers and found to be a reasonably successful method of improving solid-liquid separation. This success prompted the investigation of chemical conditioning of coal mine drainage sludge.

After a thorough search of the technical literature, it was concluded that chemical conditioning with flocculants and filter aids might be promising. Flocculants are generally high molecular weight, water soluble synthetic polymers that improve solid-liquid separation by a combination of several mechanisms such as surface charge reduction and absorption.

Charge reduction or neutralization occurs when electrical charges on the surface of the solid particles are reduced. The solid particles are then no longer repellent and can adhere or coagulate when they come in contact with each other.

Absorption is achieved when the polymer molecules attract and hold particulate matter. The polymers may also attach to each other in a bridging action which can result in the formation of clumps of particles many times larger than the original particles.

The result of charge reduction and absorption is flocculation which produces heavier, faster settling particles thereby improving solids-liquid separation.

Filter aids are chemicals which increase the permeability of a filter cake (a relatively compact sludge layer on a dewatering apparatus such as a vacuum filter). Filter aids generally increase permeability by physically disrupting the packing of the sludge as the filter cake forms, providing more channels or openings through which water may pass.

Recognizing the versatility of these types of conditioners, two different, but related studies were conducted. The major study dealt with the application of flocculants and filter aids as an aid to filtration operations. The second study dealt with the application of the same flocculants as an aid to clarification.

Flocculants were evaluated primarily as they applied to dewatering; however, these conditioners could be applied to a slurry before it entered the clarifier and simultaneously enhance both settling rates and dewaterability. Accordingly, a flocculant was selected for each sludge in terms of improved dewaterability. The same flocculant was then evaluted for any secondary benefits such as increased settling rates.

Equipment

Sludge conditioning investigations emphasizing filtration applications were conducted using a Buchner funnel filtrate recovery test. The Buchner funnel test was used because it is a commonly accepted laboratory procedure that simulates vacuum filtration operations. An illustration of the Buchner funnel apparatus is presented in Figure 12. The equipment used for the tests was an 11 centimeter Buchner funnel, Whatman Number 5 filter paper, two 1000 milliliter vacuum flasks, a vacuum gauge and a small vacuum pump. The two vacuum flasks were connected in series to the vacuum pump, the one closest to the pump fitted with the vacuum gauge and the other holding the Buchner funnel. A glass "T" was placed in the hose between the vacuum gauge and the pump with a short piece of tubing and a pinch clamp attached to the leg of the "T". The pinch clamp was used as a relief valve to control the vacuum.

The settling rate (clarification) tests were performed using two 1000 milliliter graduated cylinders and two 1000 milliliter volumetric flasks.

Filtration Procedure

The initial step in performing a Buchner funnel filtrate recovery test was preparation of the flocculant sample. Each flocculant examined was mixed with distilled water to obtain a 0.25 weight percent stock solution. The proper amount of stock solution to give the desired flocculant concentration was volumetrically added to a graduated cylinder and diluted with distilled water to a volume of 30 milliliters. In this manner the percent solids concentration of the sludge was always altered by a fixed amount for each test and variations in filtrate recovery were not due to a change in solids content.

The 30 milliliter flocculant solution was placed in one beaker and 100 milliliters of the sludge to be tested was placed in another beaker. The contents of the beakers were mixed by pouring the flocculant into the sludge and then pouring the mixture from one beaker to the other five times. The degree of mixing was thereby held constant.

A piece of filter paper was placed in the Buchner funnel, wetted and vacuum applied to seal the paper to the funnel. The vacuum was then released, the flask emptied and the conditioned sludge poured into the funnel. The vacuum was applied for one minute, released and the

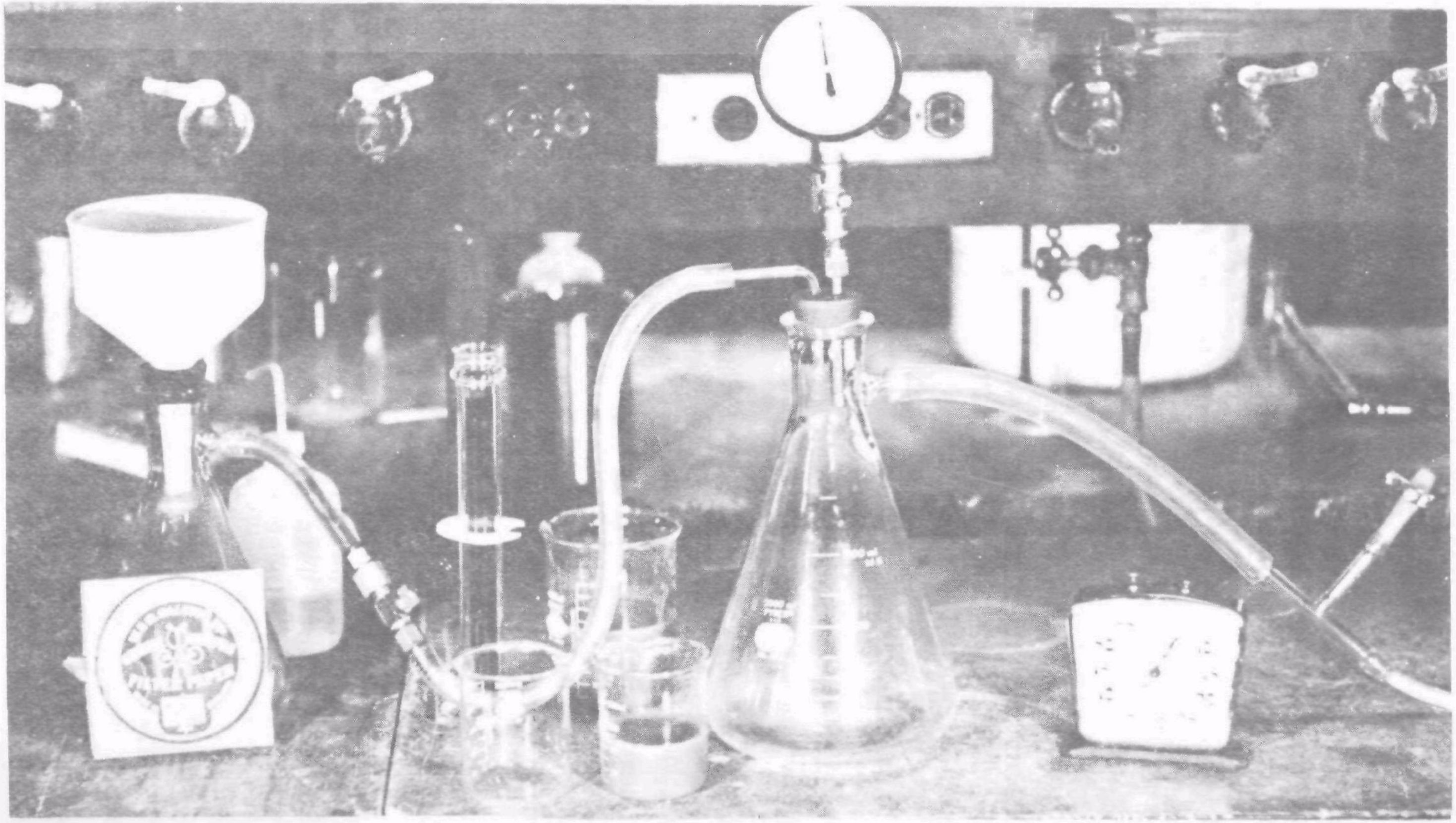


Figure 12 - BUCHNER FUNNEL APPARATUS

amount of filtrate recovered was measured and recorded.

For control purposes, the untreated sludge was tested using 30 mls of distilled water and the filtrate recovery was compared to the filtrate recovery results from the same test using conditioned sludge.

A total of 62 flocculants were examined in the first series of tests in conjunction with a lime sludge using the Buchner funnel test. Each flocculant was tested at various concentrations ranging from approximately 1 ppm to 115 ppm. The increase in filtrate recovery and optimum concentrations were recorded for a large number of cationic, anionic and non-ionic flocculants.

Filter aids (materials like diatomaceous earth which act to improve the porosity of a filter cake) were studied in a similar manner except that the desired amount of dry filter aid was poured into a graduated cylinder and distilled water was added to create a total volume of 30 milliliters. The 30 milliliter filter aid-water mixture was then mixed with 100 milliliters of the sludge. The sludge-filter aid mixture was then examined against an untreated sludge using the Buchner funnel test.

Six different filter aids were used on Norton sludge (a limestone treatment sludge) and the results examined.

Following the first series of tests, the best flocculants and/or filter aids were selected using a 30 percent or greater increase in filtrate recovery over the control filtrate recovery as a criteria. Using the Buchner funnel test again these selected flocculants were then tested against the four characteristic mine drainage sludges. From these tests the best conditioner and its optimum concentration was determined for each sludge.

Settling Rate (Clarification)

Sludge conditioning studies using flocculants were carried further into the area of clarification. Settling rate tests were conducted at the respective treatment plants using the flocculant at its optimum concentration found for that sludge from the previous Buchner funnel tests.

Two 1000 milliliter volumetric flasks were prepared, one having a 100 milliliter flocculant solution in it and the other 100 milliliters of distilled water. Both flasks were then filled with 900 milliliters of sludge slurry collected just before it entered the settling lagoon or clarifier. The two flasks were emptied into 1000 milliliter graduated cylinders and the timing clock started. As the sludges settled, interface readings were taken at various times during the test along with remarks as to the relative clarity of the water above the interface. In the case of the Norton slurry, this procedure could not be followed as a definite interface did not form. On those occasions the only data taken was a relative clarity, with and without flocculants, at various time intervals.

Filtration Results

Based upon the results of the initial screening test presented in Table 7, 14 flocculants were found to give an increase in filtrate recovered of 30 percent or greater. These 14 selected flocculants were all anionic and had a medium to high atomic weight relative to the other flocculants tested. The results of the initial screening test for filter aids is shown in Table 8. None of the filter aids gave an increase in filtrate recovered of over 30 percent. Filter aids were thereby eliminated from further study.

The results of the detailed evaluation of the 14 selected flocculants are shown in Tables 9 through 12. Shannopin sludge responded best to the use of flocculants; however, it required a relatively large dosage for maximum effect. Norton sludge on the other hand responded the least to the use of flocculants although maximum effect was obtained at a relatively small dosage. From this latter series of tests a flocculant for each of the four sludges was selected on the basis of greatest increase in filtrate recovered for that sludge. Each sludge was conditioned with its best flocculant in later dewatering studies. The selected flocculants are listed below:

1. Shannopin Sludge - Nalcolyte 673 at 111.0 ppm
2. Banning Sludge - Coagulant 2350 at 37.0 ppm
3. Edgell Sludge - Hercofloc 831 at 111.0 ppm
4. Norton Sludge - Decolyte 940 at 37.0 ppm

Clarification Results

The settling curves for Shannopin, Banning and Edgell sludges, all of which settle with a distinct interface, are shown on Figures 13 through 15. The settling curves for sludges without flocculant may differ somewhat from those shown in Figures 5 through 7 due to variances in the water treated at the same treatment plant on different days. The variance in the water treated causes a variance in the sludge produced and therefore alters the settling rates.

All of the sludges settled faster with a flocculant and all showed some increase in water clarity; however, the improvement is probably not enough to justify their use for clarification alone. In interpreting these settling curves it must be remembered that these flocculants were selected with increased filtration rates as a criteria rather than increased settling rates.

Norton sludge was tested with a flocculant and while no interface could be detected the sludge did not appear to settle faster with a flocculant and did not produce a clearer water.

Table 7

Effect of Flocculants on Filtrate Increase
Lime Sludges

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.25/lb.	*Allied Colloids, Inc. Percol 139	anionic	11.1	39.7	$0.12
$1.25/lb.	Allied Colloids, Inc. Percol 140	cationic	2.9	5.1	$0.03
$1.25/lb.	Allied Colloids, Inc. Percol 155	anionic	11.1	23.2	$0.12
$1.25/lb.	*Allied Collids, Inc. Percol 156	anionic	18.5	63.2	$0.19
$1.35/lb.	Allied Colloids, Inc. Percol 292	cationic	5.8	1.6	$0.06
$1.25/lb.	Allied Colloids, Inc. Percol 351	nonionic	3.9	7.3	$0.04
$0.24/lb.	American Cyanamid Co. Magnifloc 521 C	cationic	111.0	9.0	$0.22

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$0.28/lb.	American Cyanamid Co. Magnifloc 560 C	cationic	---	0.0	---
**NA	American Cyanamid Co. Magnifloc 570 C	cationic	3.7	0.5	NA
**NA	American Cyanamid Co. Magnifloc 571 C	cationic	37.0	3.4	NA
$1.35/lb.	American Cyanamid Co. Magnifloc 835 A	anionic	18.5	20.1	$0.21
$1.35/lb.	American Cyanamid Co. Magnifloc 836 A	anionic	3.7	25.8	$0.04
$1.50/lb.	American Cyanamid Co. Magnifloc 837 A	anionic	1.9	23.0	$0.02
$1.50/lb.	American Cyanamid Co. Magnifloc 905 N	nonionic	---	0.0	---
$1.25/lb.	American Cyanamid Co. Superfloc 16	nonionic	2.9	1.6	$0.03
$1.25/lb.	American Cyanamid Co. Superfloc 20	nonionic	2.9	4.2	$0.03

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.30/lb.	American Cyanamid Co. Superfloc 84	nonionic	3.9	6.9	$0.04
$1.30/lb.	American Cyanamid Co. Superfloc 127	nonionic	1.9	5.7	$0.02
$1.25/lb.	American Cyanamid Co. Superfloc 202	anionic	1.9	21.7	$0.02
$0.40/lb.	American Cyanamid Co. Superfloc 310	cationic	1.9	1.8	$0.01
$1.65/lb.	*Calgon Corp. Calgon 240	anionic	25.9	68.3	$0.04
$2.25/lb.	Calgon Corp. Coagulant 2256	cationic	5.8	0.5	$0.11
$2.25/lb.	Calgon Corp. Coagulant 2260	cationic	5.8	3.8	$0.11
$1.92/lb.	Calgon Corp. Coagulant 2300	nonionic	5.8	1.8	$0.09
$1.92/lb.	Calgon Corp. Coagulant 2325	anionic	3.9	3.7	$0.06

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.92/lb.	*Calgon Corp. Coagulant 2350	anionic	55.5	37.0	$0.09
$1.92/lb.	*Calgon Corp. Coagulant 2400	anionic	18.5	41.9	$0.30
$1.92/lb.	*Calgon Corp. Coagulant 2425	anionic	111.0	50.0	$1.78
**NA	Diamond Shamrock Corp. Decolyte 710	cationic	---	0.0	---
**NA	Diamond Shamrock Corp. Decolyte 720	cationic	3.7	3.6	NA
$1.41/lb.	*Diamond Shamrock Corp. Decolyte 930	anionic	1.9	32.9	$0.04
**NA	*Diamond Shamrock Corp. Decolyte 940	anionic	3.7	45.3	NA
**NA	Diamond Shamrock Corp. Decolyte 950	anionic	3.7	28.0	NA
**NA	*Dow Chemical Co. Purifloc A-21	anionic	111.0	40.4	NA

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.30/lb.	Dow Chemical Co. Separan AP 30	anionic	37.0	23.3	$0.40
$1.35/lb.	*Dow Chemical Co. Separan AP 273	anionic	11.1	31.3	$0.13
$0.45/lb.	Drew Chemical Corp. Drewfloc 3	anionic	1.9	12.0	$0.01
$0.45/lb.	Drew Chemical Corp. Drewfloc 3	cationic	---	0.0	---
$1.50/lb.	Drew Chemical Corp. Drewfloc 6	anionic	0.4	9.8	$0.01
$1.50/lb.	Drew Chemical Corp. Drewfloc 6	cationic	5.8	5.7	$0.07
$1.50/lb.	Drew Chemical Corp. Drewfloc 230	anionic	2.9	8.0	$0.04
*NA	Drew Chemical Corp. Drewfloc 230	cationic	1.9	5.7	NA
$0.30/lb.	Drew Chemical Corp. Drewfloc 410	anionic	0.6	10.2	$0.01

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.35/lb.	Hercules, Inc. Hercofloc 812	cationic	2.9	1.9	$0.03
$1.50/lb.	Hercules, Inc. Hercofloc 815	cationic	5.8	12.7	$0.07
$1.25/lb.	*Hercules, Inc. Hercofloc 818	anionic	11.1	57.4	$0.12
$1.25/lb.	Hercules, Inc. Hercofloc 821	anionic	5.8	21.2	$0.06
$1.25/lb.	Hercules, Inc. Hercofloc 827	nonionic	1.0	4.5	$0.01
$1.25/lb.	*Hercules, Inc. Hercofloc 831	anionic	18.5	60.3	$0.19
$1.25/lb.	Hercules, Inc. Hercofloc 834	cationic	1.9	12.5	$0.02
$0.49/lb.	Nalco Chemical Co. Nalcolyte 110 A	nonionic	1.0	3.2	$0.01
$0.31/lb.	Nalco Chemical Co. Nalcolyte 603	cationic	0.2	3.3	$0.01

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.37/lb.	Nalco Chemical Co. Nalcolyte 670	nonionic	---	0.0	---
$2.06/lb.	Nalco Chemical Co. Nalcolyte 672	anionic	1.9	11.4	$0.03
$1.76/lb.	*Nalco Chemical Co. Nalcolyte 673	anionic	18.5	41.7	$0.27
$1.73/lb.	Nalco Chemical Co. Nalcolyte 675 H	anionic	5.8	19.4	$0.08
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc S	cationic	1.9	6.8	$0.01
$0.11/lb.	*Narvon Mining and Chemical Co. Zeta Floc WA	anionic	5.8	31.4	$0.01
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc WA3	anionic	1.9	13.7	$0.01
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc WA5	anionic	1.9	19.0	$0.01

Table 7 (Continued)

Unit Cost	Flocculant Tested	Flocculant Type	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$17.75/100	National Starch and Chemical Co. Natron 86	cationic	---	0.0	---
$0.18/lb.	Rohm and Hass Co. Primafloc A 10	anionic	5.8	26.6	$0.01

Average increase in recovery for all flocculants - 17.3%
Average increase in recovery with anionics - 30.0%
Average increase in recovery with cationics - 3.9%
Average increase in recovery with nonionics - 3.5%

* Selected for further study.

** This flocculant no longer manufactured, price not available.

Table 8

Effect of Filter Aids on Filtrate Increase
Norton Treatment Plant Sludge
(Average Percent Solids 4.1)

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Material Used As Filter Aid	Optimum Concentration (ppm)	Filtrate Increase With Filter Aid (percent)	Flocculant Cost Per 1000 Gallons
$0.50/ton	Municipal Incinerator Flyash	---	0.0	---
$50.80/ton	Bituminous Coal crushed to -60 mesh	3,850	6.2	$0.82
$160.00/ton	Johns-Manville Hyflo Super-Cell	3,850	8.4	$2.57
$160.00/ton	Johns-Manville Celite 501	3,850	10.8	$2.57
$160.00/ton	Johns-Manville Celite 503	3,850	3.2	$2.57
$0.50/ton	Wet Collected limestone modified flyash	40,000	2.7	$0.08

Table 9

Effect of Selected Flocculants on Filtrate Increase
Shannopin Treatment Plant Sludge
(Average Percent Solids 5.3)

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.25/lb.	Allied Colloids, Inc. Percol 139	55.5	56.5	$0.58
$1.25/lb.	Allied Colloids, Inc. Percol 156	37.0	54.2	$0.39
$1.65/lb.	Calgon Corp. Calgon 240	74.0	58.3	$1.02
$1.92/lb.	Calgon Corp. Coagulant 2350	44.4	61.7	$0.71
$1.92/lb.	Calgon Corp. Coagulant 2400	44.4	57.4	$0.71
$1.92/lb.	Calgon Corp. Coagulant 2425	44.4	56.0	$0.71
$1.41/lb.	Diamond Shamrock Corp. Decolyte 930	111.0	10.1	$1.31

Table 9 (Continued)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
**NA	Diamond Shamrock Corp. Decolyte 940	111.0	32.1	NA
**NA	Dow Chemical Co. Purifloc A 21	111.0	58.7	NA
$1.35/lb.	Dow Chemical Co. Separan AP 273	55.5	60.0	$0.63
$1.25/lb.	Hercules, Inc. Hercofloc 818	44.4	50.0	$0.46
$1.25/lb.	Hercules, Inc. Hercofloc 831	74.0	65.9	$0.77
$1.76/lb.	Nalco Chemical Co. Nalcolyte 673	111.0	13.7	$1.63
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc WA	111.0	13.7	$0.10
	Average	73.5	50.1	

** This flocculant no longer manufactured, price not available.

Table 10

Effect of Selected Flocculants on Filtrate Increase
Banning Treatment Plant Sludge
(Average Percent Solids 1.2)

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.25/lb	Allied Colloids, Inc. Percol 139	11.1	19.3	$0.12
$1.25/lb.	Allied Colloids, Inc. Percol 156	25.9	25.3	$0.27
$1.65/lb.	Calgon Corp. Calgon 240	25.9	23.7	$0.36
$1.92/lb.	Calgon Corp. Coagulant 2350	37.0	42.8	$0.59
$1.92/lb.	Calgon Corp. Coagulant 2400	37.0	32.9	$0.59
$1.92/lb.	Calgon Corp. Coagulant 2425	18.5	30.3	$0.30
$1.41/lb.	Diamond Shamrock Corp. Decolyte 930	1.9	32.9	$0.02

Table 10 (Continued)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
**NA	Diamond Shamrock Corp. Decolyte 940	3.7	41.1	NA
**NA	Dow Chemical Co. Purifloc A 21	37.0	25.6	NA
$1.35/lb.	Dow Chemical Co. Separan AP 273	11.1	31.3	$0.12
$1.25/lb.	Hercules, Inc. Hercofloc 818	11.1	19.0	$0.12
$1.25/lb.	Hercules, Inc. Hercofloc 831	25.9	37.1	$0.27
$1.76/lb.	Nalco Chemical Co. Nalcolyte 673	111.0	26.7	$1.63
$0.11/ln.	Narvon Mining and Chemical Co. Zeta Floc WA	37.0	33.8	$0.03
	Average	28.2	30.1	

** This flocculant no longer manufactured, price not available.

Table 11

Effect of Selected Flocculants on Filtrate Increase
Norton Treatment Plant Sludge
(Average Percent Solids 4.1)

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.25/lb.	Allied Colloids, Inc. Percol 139	3.7	15.7	$0.04
$1.25/lb.	Allied Colloids, Inc. Percol 156	9.3	12.6	$0.10
$1.65/lb.	Calgon Corp. Calgon 240	55.5	12.7	$0.76
$1.92/lb.	Calgon Corp. Coagulant 2350	11.1	6.5	$0.18
$1.92/lb.	Calgon Corp. Coagulant 2400	18.5	14.9	$0.30
$1.92/lb.	Calgon Corp. Coagulant 2425	7.4	18.4	$0.12
$1.41/lb.	Diamond Shamrock Corp. Decolyte 930	25.9	26.1	$0.30

Table 11 (Continued)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
**NA	Diamond Shamrock Corp. Decolyte 940	37.0	31.7	NA
**NA	Dow Chemical Co. Purifloc A 21	55.5	25.6	NA
$1.35/lb.	Dow Chemical Co. Separan AP 273	37.0	13.5	$0.42
$1.25/lb.	Hercules, Inc. Hercofloc 818	37.0	20.8	$0.39
$1.25/lb.	Hercules, Inc. Hercofloc 831	37.0	13.5	$0.39
$1.76/lb.	Nalco Chemical Co. Nalcolyte 673	37.0	17.3	$0.54
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc WA	111.0	12.2	$0.10
	Average	34.5	17.3	

** This flocculant no longer manufactured, price not available.

Table 12

Effect of Selected Flocculants on Filtrate Increase
Edgell Treatment Plant Sludge
(Average Percent Solids 3.6)

(Unit costs based on purchase quantity of 1,000 pounds, F.O.B.)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
$1.25/lb.	Allied Colloids, Inc. Percol 139	25.9	16.7	$0.27
$1.25/lb.	Allied Colloids, Inc. Percol 156	92.5	42.2	$0.96
$1.65/lb.	Calgon Corp. Calgon 240	37.0	67.8	$0.51
$1.92/lb.	Calgon Corp. Coagulant 2350	37.0	50.0	$0.59
$1.92/lb.	Calgon Corp. Coagulant 2400	92.5	35.4	$1.48
$1.92/lb.	Calgon Corp. Coagulant 2425	74.0	55.8	$1.19
$1.41/lb.	Diamond Shamrock Corp. Decolyte 930	111.0	19.8	$1.30

Table 12 (Continued)

Unit Cost	Flocculant Tested	Optimum Flocculant Concentration (ppm)	Filtrate Increase With Flocculant (percent)	Flocculant Cost Per 1000 Gallons
**NA	Diamond Shamrock Corp. Decolyte 940	111.0	24.0	NA
**NA	Dow Chemical Co. Purifloc A 21	92.5	22.2	NA
$1.35/lb.	Dow Chemical Co. Separan AP 273	81.4	55.8	$0.92
$1.25/lb.	Hercules, Inc. Hercofloc 818	111.0	50.0	$1.15
$1.25/lb.	Hercules, Inc. Hercofloc 831	111.0	73.8	$1.15
$1.76/lb.	Nalco Chemical Co. Nalcolyte 673	111.0	53.3	$1.63
$0.11/lb.	Narvon Mining and Chemical Co. Zeta Floc WA	111.0	4.5	$0.10
	Average	85.6	40.8	

**
This flocculant no longer manufactured, price no available.

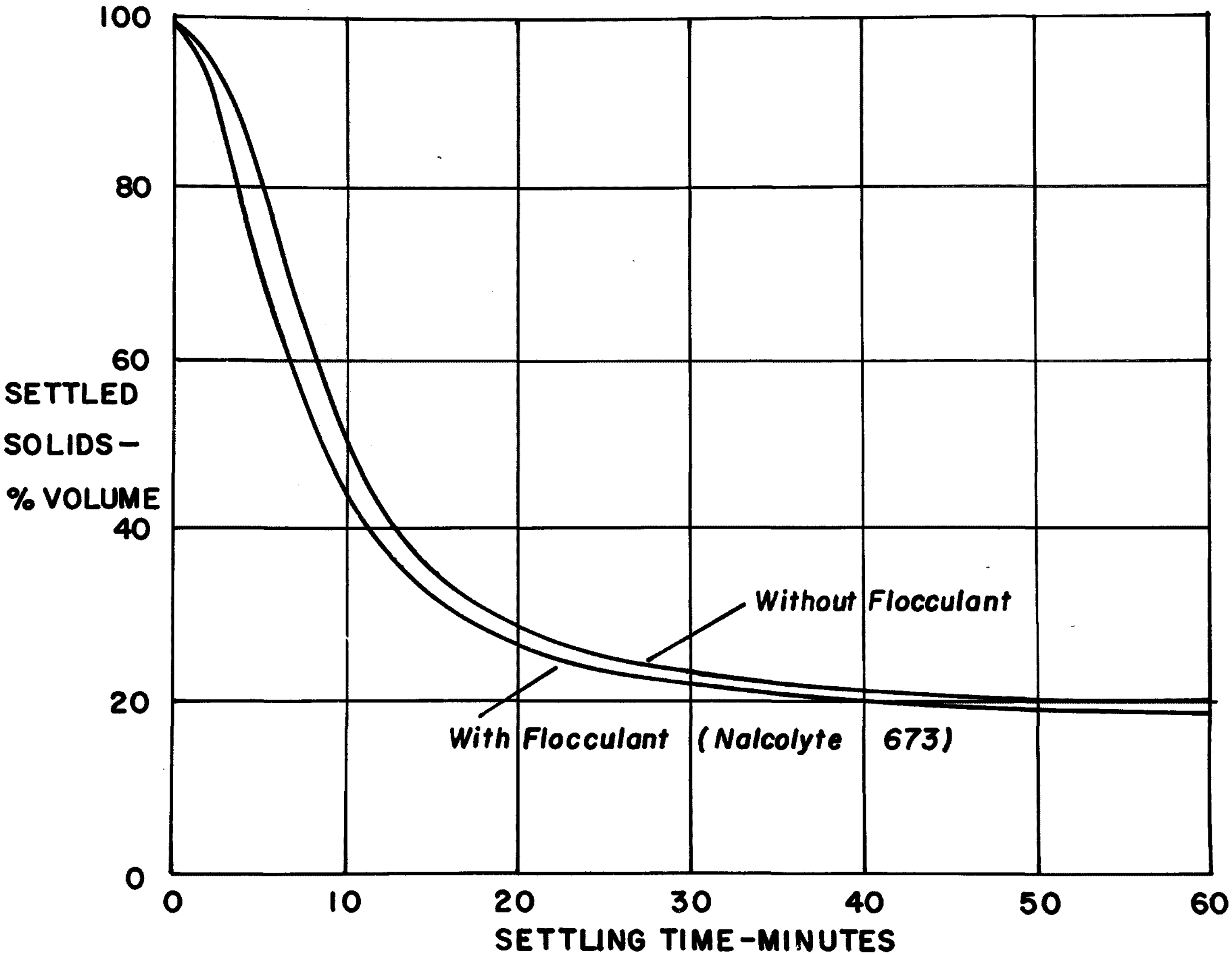


Figure 13 – SETTLING RATE OF CONDITIONED MIXING TANK SLUDGE FROM SHANNOPIN TREATMENT PLANT

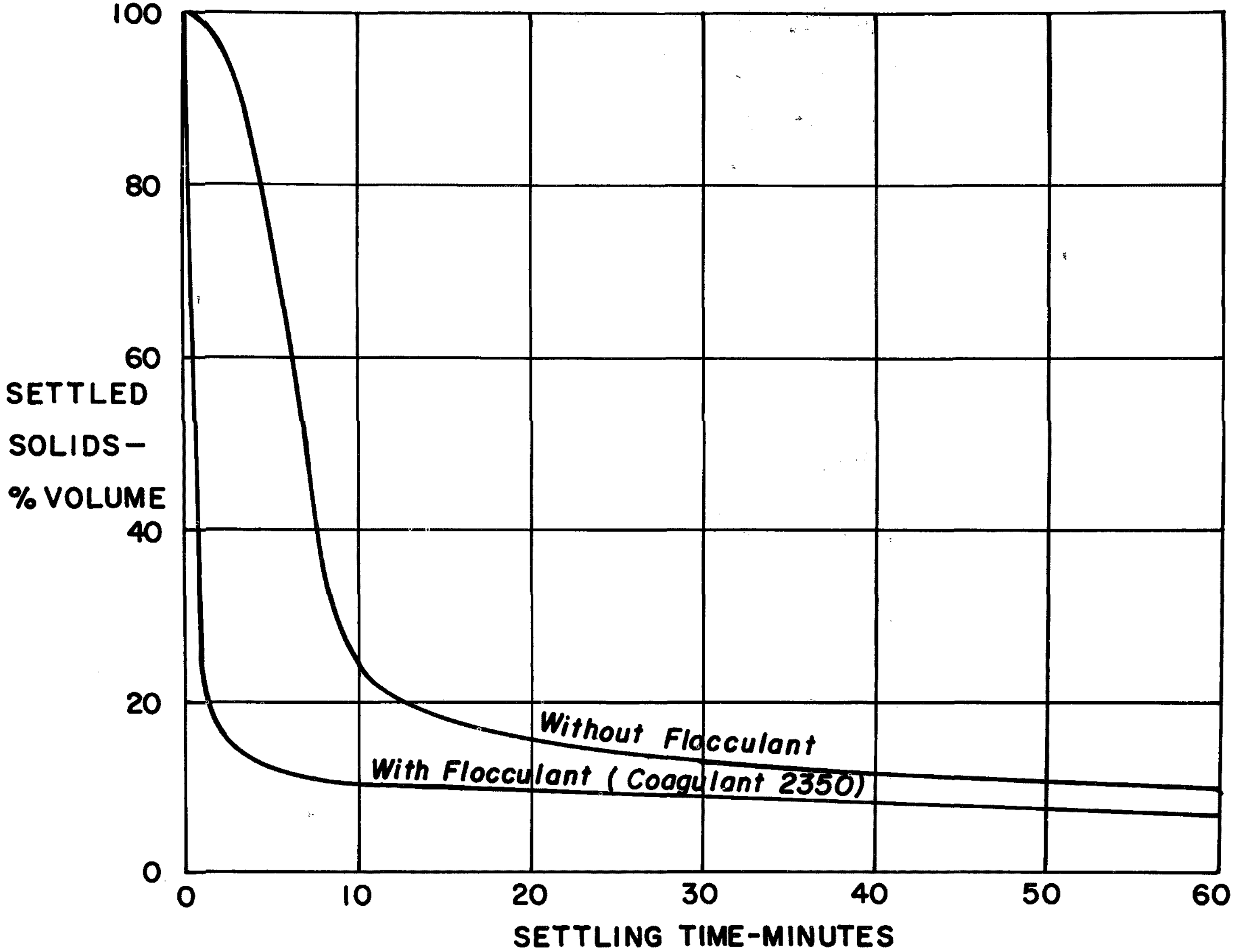


Figure 14 — SETTLING RATE OF CONDITIONED AERATOR SLUDGE FROM BANNING TREATMENT PLANT

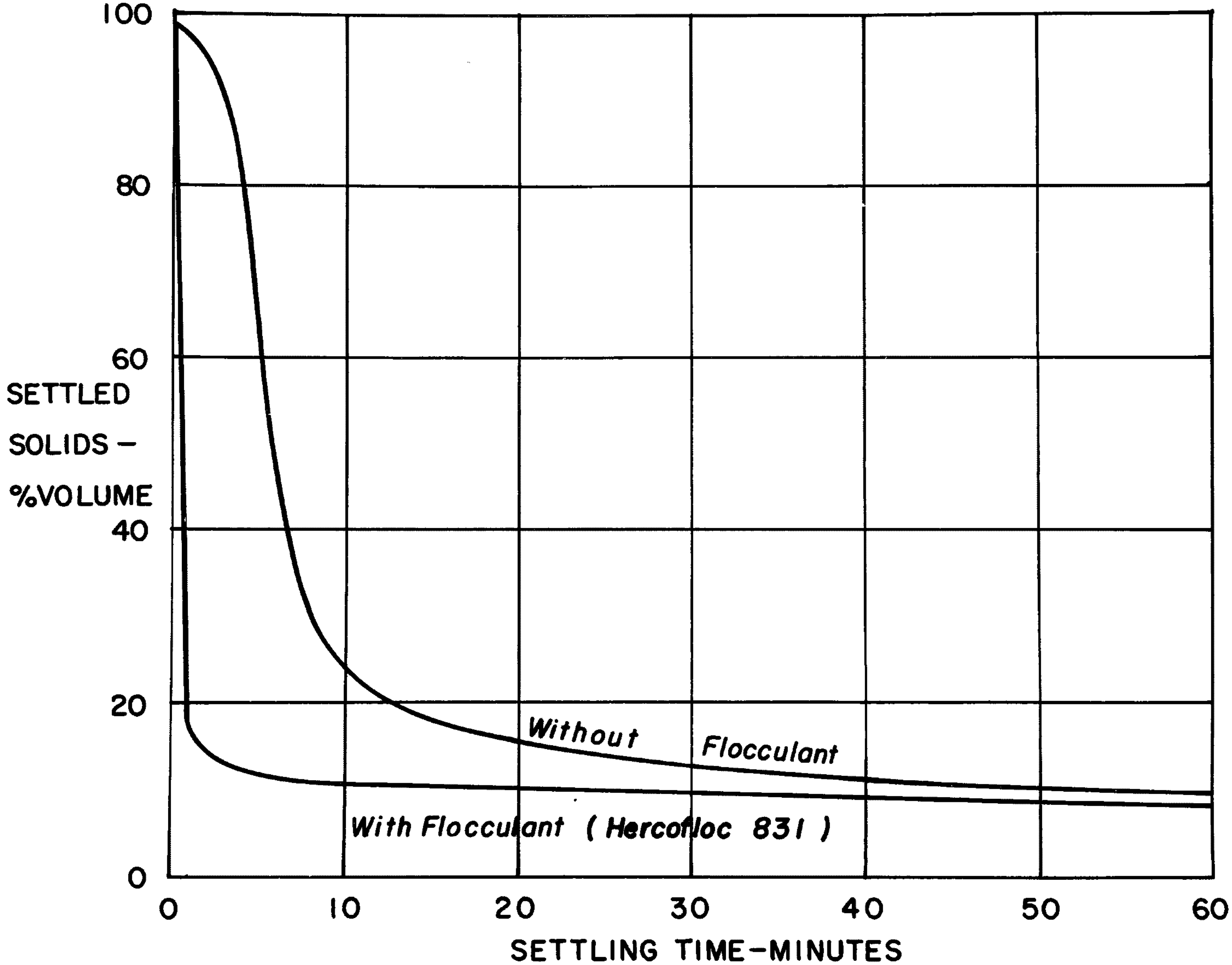


Figure 15 – SETTLING RATE OF CONDITIONED AERATOR SLUDGE FROM EDGELL TREATMENT PLANT

SECTION VI

SLUDGE DEWATERING

The primary objective of this research was to evaluate various dewatering systems as they apply to the sludge produced by the neutralization of acid mine drainage. The systems evaluated were:

1. Conventional Rotary Vacuum Filtration
2. Rotary Precoat Vacuum Filtration
3. Pressure Filtration
4. Porous Bed Filtration
5. Thermal Spray Drying
6. Centrifugation

The evaluation of each of the above dewatering systems is presented in four parts. The first part is a brief discussion of the dewatering system as used in industry. This section includes a brief description of a commercial size unit of the dewatering system and its general method of operation. The second part is a description of the test apparatus used. Generally the test equipment was bench scale; however, some of the dewatering systems were studied on a semi-pilot plant scale. The procedure used in making the dewatering tests is described in the third section. And the fourth section is a discussion of the results of the bench scale tests. The results section is intended to show whether it is technically feasible to dewater each of the four sludges by the method in question and to compare the relative efficiency of this method on each sludge.

The results obtained with these different systems vary with changes in the solids content of the sludge to be dewatered. Also, the sludge collected from the settling lagoons of Shannopin, Norton and Edgell varied in solids content with each field trip. Since the sludge was pumped from below several feet of water varying amounts of water would be pulled into the pump with the sludge. Banning sludge was collected as underflow from the clarifier and showed no significant change in solids content, averaging very close to 0.5 percent solids. In order to maintain a fair basis for comparison of the various dewatering systems it was necessary to assign a solids content for each sludge as follows: Shannopin sludge 2.1 percent, Edgell sludge 2.7 percent and Norton sludge 8.0 percent. In some cases additional work was done at a higher or lower solids content than that specified in order to determine the effect of a variance in solids content on a specific dewatering system. If the solids content of a sludge as collected was found to be less than the desired solids content, it was allowed to settle and some water decanted until the correct solids content was obtained. In cases where the solids content as collected was too high, treated water from the plant in question was added until the solids content was correct.

Conventional Vacuum Filtration

General

A conventional rotary vacuum filtration unit consists of a horizontal drum covered with a filter media. Vacuum is applied to segments of the filter media from within the drum as the drum is rotated. As the filtrate is pulled through the filter media the solids are deposited on the media.

Rotary vacuum filtration may be divided into three phases. The first phase is filtration which occurs on a portion of the drum during the time when that portion of the drum is submerged in the sludge. As a segment of the drum rotates into the sludge, vacuum is applied and filtrate is drawn through the filter media and discharged. Concurrently, sludge solids are deposited on the filter media face to form a cake. As the sludge cake becomes thicker, its resistance to the passage of filtrate increases. The drum speed and amount of submergence of the drum are adjusted such that each segment of the drum leaves the slurry before the increasing cake resistance reduces the filtrate flow rate below an acceptable level.

The second phase of the operation occurs during the time a segment of the drum leaves the sludge and before the cake is removed. It is during this time that the cake is dried. As the drum leaves the sludge the cake is still under vacuum and additional moisture contained within the cake is drawn out. This phase may be of varying length depending on the desired dryness of the cake and the time required to achieve this dryness. The cake has reached its highest practical degree of dryness when it cracks as air will then be drawn through the cracks in the cake rather than through the cake itself.

The third phase, cake removal, begins after the cake has reached acceptable dryness. At this point the vacuum is removed and the cake is discharged. The most common types of conventional cake discharge devices are a scraper discharge, a wire discharge or a string discharge. The scraper, the most commonly used type, is a sharp blade which is mounted on the side of the filter tank and scrapes the filter cake from the drum. However, when thin sticky cakes are encountered, a taut longitudinal wire may be used. The longitudinal wire is under great tension and is fixed in the same position as the scraper tip; this allows the cake to be peeled off the drum. When cakes which adhere strongly to the filter cloth are encountered a string type discharge may be used. With the string type discharge, annular strings are spaced about 3/8 inch apart and are around the drum and on top of the filter cloth. The strings are led from the discharge point on the drum to a small roller away from the drum and then back to the drum again. The cake is supported by these strings. As the strings leave the drum the cake is led off the drum and is discharged at the sharp bend made by the roller in returning the string to the drum.[10] All of the operations described are of a continuous nature so all three phases are occurring simultaneously on different portions of the drums surface.

Test Apparatus - Description

Bench scale experiments were performed with a 0.1 square foot surface area, Dorr-Oliver filter leaf apparatus shown in Figure 16. This leaf was designed to simulate the function of an equal area on the surface of a full size conventional rotary vacuum filtration drum.

The filter leaf was connected to two graduated receivers by hoses and a two way valve. The vacuum source was connected to the receivers in a like manner so that one receiver could be collecting filtrate while the other was being vented and drained. The vacuum source consisted of a small vacuum pump which was connected by a hose to a 1000 milliliter volumetric flask. The flask was equipped with a vacuum gage which registered inches of mercury and an air bleed to reduce vacuum to the leaf. The purpose of the flask was to keep water from being pulled into the vacuum pump. The test apparatus also included a four liter container which held the sludge to be tested. The sludge was stirred manually with a glass rod in order to keep the sludge particles in suspension.

Various filter cloths from Eimco Filter Media Corporation and National Filter Media Corporation were used in these tests.

Procedure

The three phases of a rotary vacuum filtration cycle were duplicated with the filter leaf. First the filter leaf was submerged in the sludge container under vacuum to duplicate the filtration phase of the filter drum's cycle. Second the filter leaf was lifted from the sludge and placed face up on a ring stand to duplicate the drying phase. The third phase involved removing the built up sludge cake by scraping it away after air had been applied behind the filter media to loosen the cake. As with rotary vacuum filtration, the filter media was under vacuum for the entire cycle except for cake discharge. The amount of vacuum used was 24 to 25 inches of Hg.

Since a minimum cake thickness of 1/8 inch was necessary to permit proper discharge of the cake from the filter cloth,[11] the conditions necessary to produce a 1/8 inch thick cake were determined. Length of the first phase of a conventional rotary vacuum filtration cycle (the submergence time of the leaf into the sludge) determines the thickness of the cake formed. By experimentation the length of time

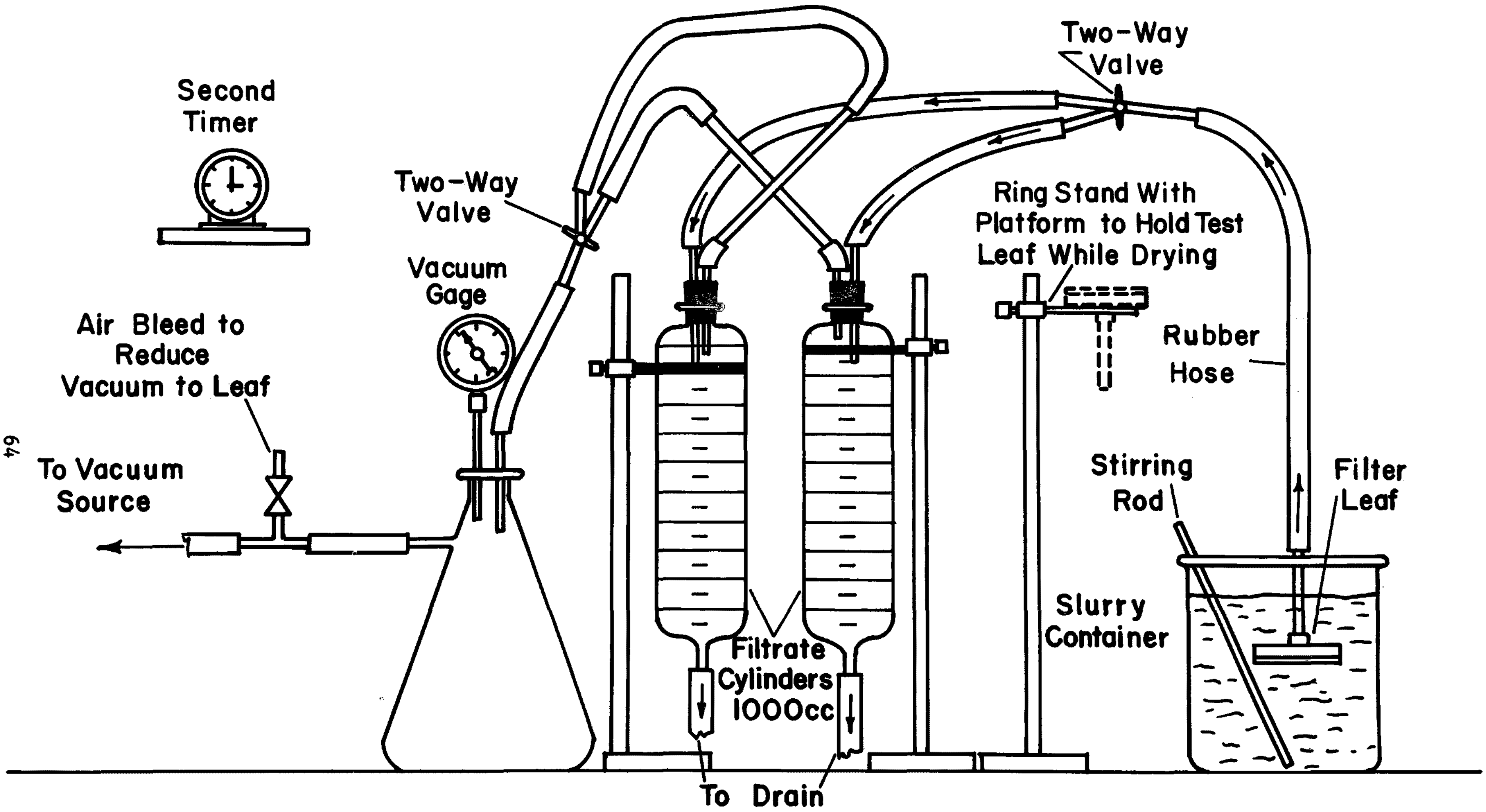


Figure 16 - FILTER LEAF APPARATUS

required for a 1/8 inch cake build up for each sludge was found. This time was the minimum submergence time for the cycle. For the second phase the leaf was placed on a ring stand and vacuum dried. The drying time was found experimentally by allowing the cake to dry until it cracked, at which time the cake had reached its highest practical degree of dryness. The third phase of the cycle (the time required to remove the cake from the filter cloth) is constant depending upon the type of machinery used. The amount of filtrate recovered each cycle and the time required for the cake to crack during drying along with any observations as to condition of the sludge cake were recorded. After the optimum cycle time was established for a sludge, the sludge cake formed was analyzed for solids content on an Ohaus moisture determination balance.

All four sludges, with and without flocculants, were studied in the same manner. The sludge was conditioned with the flocculant that was found in the earlier Buchner funnel tests to give the highest increase in filtration rates.

Results

Results obtained with conventional rotary vacuum filtration were good for Norton sludge but poor for Banning, Edgell and Shannopin sludges.

In the tests 38 filter cloths from Eimco Corporation and 5 filter cloths from National Corporation were investigated with a lime neutralized sludge. Of these 43 different cloths four were chosen according to best filtrate clarity and best filtration rates obtained. These four cloths were all from Eimco Corporation and were NY-301, CO-3, NY-518F, and Popr-913F. These cloths were tested with each of the four sludges both with and without the optimum concentration of flocculant added as determined by Buchner funnel tests. It was found that filter cloth CO-3 gave the best results for Banning, Edgell and Norton sludges and cloth NY-301 gave the best results for Shannopin sludge. Results on the four sludges and their respective filter cloths are shown in Tables 13 through 16.

A minimum sludge cake build up of 1/8 inch is necessary to allow proper discharge of the cake from the filter cloth. Vacuum filters are normally designed with a variable speed drive to operate at between 1 to 10 minutes per revolution (mpr). This would mean that cake formation time for a 1/3 submergence drum cannot exceed 3 1/3 minutes. A drum with 1/2 or greater submergence was not considered due to high capital and operating costs. If a cake thickness of 1/8 inch or greater can be formed in this time, a conventional vacuum filter should be considered. If the cake thick-

Table 13

Conventional Vacuum Filtration
Shannopin Treatment Plant Sludge

Filter Cloth - Eimco NY 301

Feed Solids (percent)	Dip Time (min)	Time to Crack (sec)	Sludge Dewatered (gal/sq ft/min)	Filtrate Quality Nonfilterable Solids (ppm)	Cake Thickness (in)	Solids (percent)
2.4	10	120	0.0576	997	0.1250	17.0
2.4*	10	17	0.0731	23	0.1250	12.6
4.2	10	30	0.0280	343	0.3750	15.1
4.0*	5	420	0.0474	25	0.1875	15.5

* With Nalcolyte 673 added (flocculant)

Table 14

Conventional Vacuum Filtration
Banning Treatment Plant Sludge

Filter Cloth - Eimco CO-3

Feed Solids (percent)	Dip Time (min)	Time to Crack (sec)	Sludge Dewatered (gal/sq ft/min)	Filtrate Quality Nonfilterable Solids (ppm)	Cake Thickness (in)	Solids (percent)
0.4	5	25	0.1237	618	0.125	11.2
0.4*	10	20	0.0845	23	0.125	9.2
2.2	5	120	0.0694	42	0.250	8.8
2.1*	5	120	0.0712	3	0.250	10.1

* With Coagulant 2350 added (flocculant)

Table 15

Conventional Vacuum Filtration
Norton Treatment Plant Sludge

Filter Cloth - Emico - CO-3

Feed Solids (percent)	Dip Time (min)	Time to Crack (sec)	Sludge Dewatered (gal/sq ft/min)	Filtrate Quality Nonfilterable Solids (ppm)	Cake Thickness (in)	Solids (percent)
8.1	20	15	0.1140	33	0.1250	19.8
8.1*	20	10	0.1910	13	0.1875	19.2
12.3	20	20	0.1660	-	0.3750	21.3
12.3*	20	20	0.1660	-	0.3750	20.3

* With Decolyte 940 (flocculant)

Table 16

Conventional Vacuum Filtration
Edgell Treatment Plant Sludge

Filter Cloth - Emico - CO-3

Feed Solids (percent)	Dip Time (min)	Time to Crack (sec)	Sludge Dewatered (gal/sq ft/min)	Filtrate Quality Nonfilterable Solids (ppm)	Cake Thickness (in)	Solids (percent)
3.0	10	70	0.0323	100	0.0625	30.9
3.0*	10	60	0.0396	58	0.125	24.2
6.8	10	300	0.0321	30	0.25	28.0
7.2*	10	260	0.0272	-	0.1875	29.0

* With Hercofloc 831 added (flocculant)

ness is less than this, then a precoat filter should be considered.[11]

Of the four representative sludges only Norton produced a dischargeable cake within the time limitation (3-1/3 minutes) imposed by a conventional rotary vacuum filter unit. It might be possible to use special equipment to handle the other three sludges by conventional vacuum filtration; however, the cost of such equipment would be high.

Norton sludge was tested at 8.1 and 12.3 percent solids both with and without flocculant conditioning. At 8.1 percent solids without flocculant the best dip or submergence time was found to be 20 seconds. This gave a cycle time of one minute at 1/3 submergence which is the minimum cycle time required for conventional rotary vacuum filtration and the minimum cake thickness of 1/8 inch was obtained.

When the flocculant was added to the 8.1 percent solids Norton sludge there was an increase in filtration rate, filtrate clarity and cake thickness. These increases were not observed when the flocculant was applied to the 12.3 percent solids Norton sludge. However, the final percent solids of the cake was not changed significantly with the addition of the flocculant to the sludge for either the 8.1 or the 12.3 percent solids Norton sludge.

The filtrate obtained from each of the four sludges using the best available filter cloth generally was not of sufficiently high clarity to warrant its discharge into a stream.

Rotary Precoat Vacuum Filtration

General

Rotary precoat vacuum filtration is similar to conventional rotary vacuum filtration with one major exception; the application of a precoat (generally diatomite) to the filter prior to actual filtration. Diatomite is the siliceous skeletal remains of single - celled aquatic plant life called diatoms. The diatoms form a permeable coating on the filter that allows the filtrate to pass through easily while sludge solids are trapped, thereby producing a filtrate of very high clarity.

The rotary precoat vacuum filtration unit, like the conventional rotary vacuum filtration unit, consists of a horizontal drum that is covered with a filter media. Vacuum is applied to segments of the filter media from within the drum and the drum is rotated. Initially the drum is immersed in a slurry of the precoat and an increasingly thick cake of diatomite is formed on the drum as the fluid is pulled through the filter media and the solids deposited on the media. After the precoat has reached sufficient thickness (several inches depending on the length of time the filter is to be continuously operated), it is shaved smooth and dewatering can now begin.

Rotary precoat vacuum filtration may be divided into three phases. The first and second phases are essentially the same as in conventional rotary vacuum filtration. The third phase, cake removal, begins when the cake has reached acceptable dryness. In practice the cake must be removed before it cracks as cracking in the cake tends to cause cracking and gouging of the precoat. Before cracking occurs the cake and a few thousandths of an inch of precoat are cut away by means of a continuously advancing knife. The sludge cake and the precoat are then discharged. The fresh surface of precoat now exposed by the cut is rotated into the sludge once more to again start the filtration phase.

Test Apparatus - Description

Bench scale experiments were performed with a 0.1 square foot surface area precoat filter leaf apparatus rented from Johns-Manville Research Center. This leaf, shown in Figure 17, was designed to simulate the function of an equal area on the surface of a full size rotary precoat vacuum filtration drum.

The filter leaf was composed of two parts: a collar and within the collar, a moveable septum that together act as a system to hold and advance the precoat.

The septum (or filter media) was circular and mounted on top of a screw shaft that moves the septum up or down within the cylindrical collar. An indicator fixed to the screw regulates, in thousands of an inch, movement of the septum within the collar. The filter leaf was connected to two graduated receivers by hoses and a two way valve. The vacuum source was connected to the receiver in such a manner that one receiver was collecting filtrate while the other was vented and being drained. This dual setup allowed tests of indefinite duration since one receiver could easily be drained as the other was filled and the flow of filtrate diverted to the freshly drained receiver with minimal effect on filtrate recovery.

To assure the stability of the apparatus the filter leaf was attached to a mounting that provided support. In addition to providing support the mounting also allowed the leaf to rotate in both the horizontal and vertical directions.

The test apparatus also included a small slurry tank that had a 1.5 liter capacity. The slurry tank was equipped with a variable speed agitator used to keep the sludge particles in suspension.

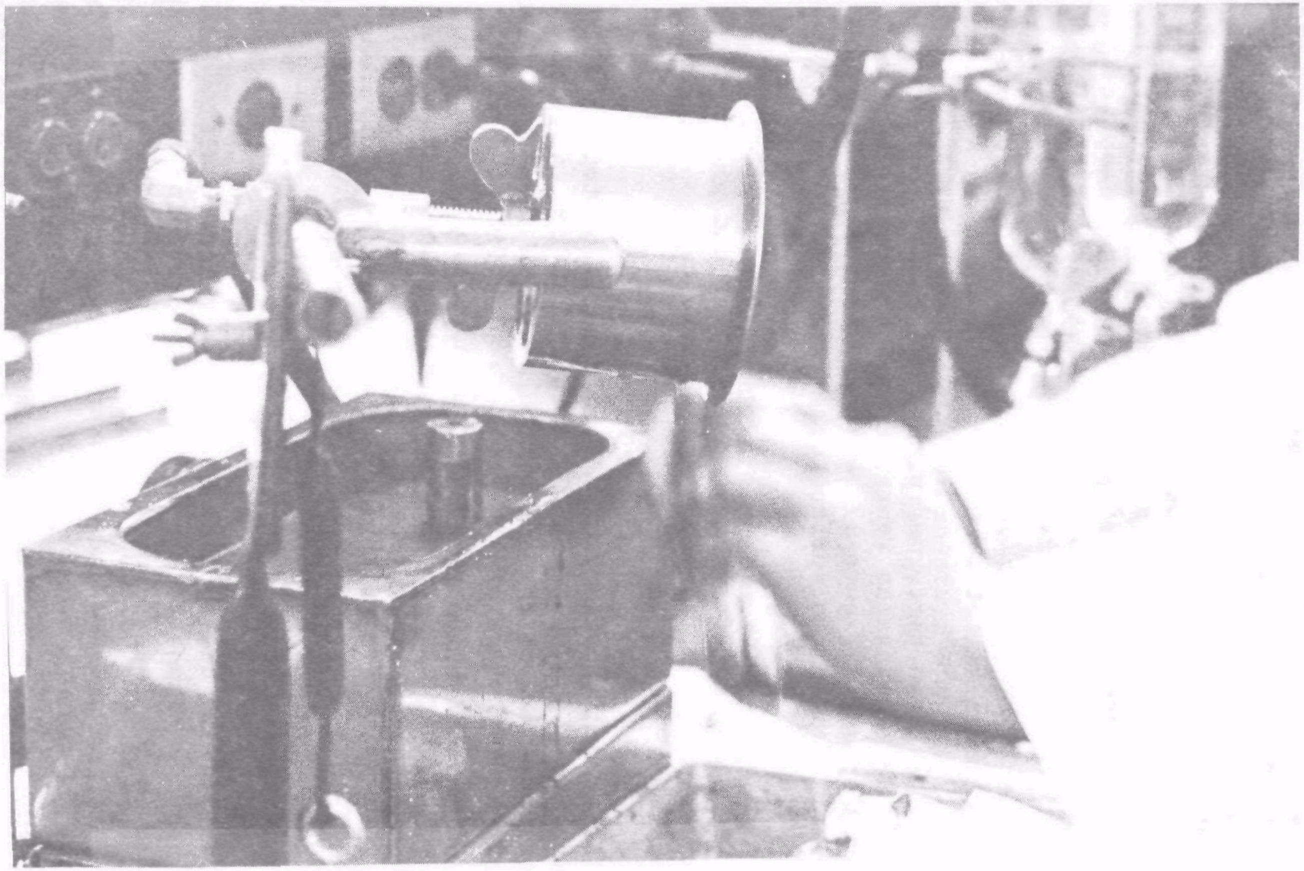


Figure 17 - PRECOAT FILTER LEAF APPARATUS

Procedure

Preparing the precoat filter leaf for experimentation was done in such a manner as to simulate the formation of a precoat on a full sized filter drum. Precoating was initiated by pouring approximately 400 mls of a 2 percent by weight precoat slurry on the filter leaf, allowing the precoat layer to filter dry and repeating the process until the desired precoat thickness was reached. Alternately adding slurry and filtering it dry produces a cake with a laminar structure similar to that which forms on a filter drum. Great care had to be taken to avoid erosion when adding slurry onto an already formed precoat. After the precoat cake was formed, shrinkage cracks were repaired by spooning the slurry onto the surface and the surface was shaved smooth.

Following the precoating operation, the three phases of a rotary precoat vacuum filtration cycle were duplicated with the filter leaf. The filter leaf was submerged in the slurry tank under vacuum to duplicate the filtration phase of a filter drum's cycle. The filter leaf was then lifted from the sludge and rotated to a vertical position to duplicate the drying phase. The third phase involved rotating the leaf to a horizontal position, advancing the cake and cutting away the sludge cake and a few mils of precoat with a sharp knife. When the cake was advanced, a portion of the precoat extended beyond the collar of the filter leaf could easily be cut away. The filter leaf was under vacuum for the entire cycle. The vacuum used was about 25 inches of Hg.

Four variables were examined during the precoat filter leaf dip tests: cycle time, depth of cut, type of precoat and flocculant effects.

A test was initiated by first determining the optimum cycle time for a sludge. The following cycle times examined, which represent submergence, drying and cutting times respectively, are considered typical for difficult to dewater sludges in industrial operations: 20-15-10 seconds, 30-20-10 seconds and 45-35-10 seconds. The cycle chosen as being best was that which produced the highest filtrate recovery and an acceptably smooth cut.

After the best cycle time was chosen, the second variable, depth of cut, was evaluated. Depth of cut refers to the thickness of precoat cut away with the sludge cake. As a sludge cake forms on the precoat there was some penetration of the precoat by the sludge particles. The depth of penetration depended on the sludge being dewatered and the type of precoat used. If the sludge particles were not removed from the precoat they would cause a decrease in permeability which is sometimes called blinding. To alleviate this blinding during each cycle the cut had to be deep enough to remove all of the particles. Starting with the 15 mil cut used during the previous cycle time the cut was made progressively shallower until filtration rates began to drop. The point at which filtration rates dropped represents the minimum usage of precoat concurrent with the maximum filtration rates.

Following the study of the best cut for the initial type of precoat, a different precoat was then evaluated. The cycle time did not change when investigating different precoats on the same sludge. Therefore, the optimum cut to allow maximum filtration was the only data taken following a change from one precoat to another.

The last variable studied was the use of a flocculant to increase filtration rates. The sludge was conditioned with the flocculant that was found by the Buchner funnel tests to give the highest increase in filtration rates. The same battery of tests (cycle time, depth of cut) were employed on the flocculant conditioned sludge and the results compared. All four sludges were studied both preconditioned and non-conditioned in the same manner using the tests described above.

The amount of filtrate recovered in each cycle was recorded along with any observations as to sludge cake or precoat condition. Since in some cases much difficulty was encountered in obtaining a smooth cut, each cut was rated on an arbitrary scale of smooth, slight gouging and gouging. When the optimum combination of the four variables for a sludge was found, the sludge cake formed by those variables and the precoat cut away with it were analyzed for solids content on the Ohaus determination moisture balance.

Results

The results obtained with precoat vacuum filtration were generally good for Edgell, Shannopin and Banning sludges and rather poor for Norton sludge.

Precoat vacuum filtration was not found to be feasible for Norton sludge as it was not possible to obtain a smooth cut. The Norton sludge would often crack the precoat and would always gouge it with the result that on successive cuts the gouged portion of the precoat face would not be cleaned of sludge. Various precoats were tried, the drying time was reduced to as little as one second, and both flocculants and body feed were added in an attempt to eliminate the gouging, but with little success. It appeared as though the Norton sludge attached itself unusually well to the precoat during filtration and as the sludge cake shrunk during drying, pieces of precoat were pulled away from the precoat cake.

Precoat vacuum filtration was found to be feasible for Edgell, Shannopin, and Banning sludges. Smooth cuts were obtained and in general it was found that these three sludges would be suitable for dewatering by precoat vacuum filtration.

The results of the precoat filter leaf tests are summarized in Table 17. The best cycle time for all three sludges was 20 seconds filtration, 15 seconds drying and 10 seconds cutting. This means a total drum speed of 45 seconds per revolution which is generally considered to be the fastest practical drum speed. A fast drum speed was necessary for efficient operation due to the rapid increase in cake resistance during filtration shown by a sharp decrease in filtrate flow rate with time. In addition a quick formation of cracks in the filter cake during drying was observed which means that only a short drying may be used since the cake must be removed before it cracks. When a short filtration time is indicated it could be achieved by a low submergence; however, since this would increase the drying time another solution had to be found. The only other mechanism available to reduce filtration time was drum speed and therefore a fast drum speed was used to provide a short filtration time and a short drying time.

The best precoat was generally Celite 501 although Hyflo Super-Cel proved best in two cases. These are both "fairly tight" precoats (capable of retaining fine particles) as would be expected since the sludge particles were small in size.

In the absence of detailed cost figures, as would be generated by a pilot plant operation, it was felt that the most realistic approach to the problem of precoat usage versus filtrate recovery rates was to maximize filtrate recovery rates. Using this criteria the optimum cut was determined to be either 4 or 5 mils per minute for all sludges tested with or without flocculants.

The highest filtrate recovery rates without a flocculant were obtained with Banning sludge which has the lowest solids content of the three sludges which were found to be suitable for dewatering by precoat vacuum filtration. The lowest recovery rates were obtained with Edgell sludge which has the highest solids content. The only unexpected development in recovery rates was the high recovery rate of Shannopin sludge with flocculants. The average recovery rate of flocculated Shannopin sludge was slightly higher than that for flocculated Edgell sludge.

The highest cake solids was obtained with Edgell sludge both with and without flocculant. In all cases the solids content of the filter cakes dropped with the use of flocculants, apparently due to the formation of a thicker (and therefore more difficult to dry) cake produced by the higher filtration rates. A strong inverse relationship may be noticed between filtration rates and cake solids content.

Table 17

Optimum Conditions for Rotary Precoat Vacuum Filtration

Sludge	Flocculant	Cycle Time Filter/Dry/Cut (seconds)	Precoat	Knife Advance Rate (mils/min)	Sludge Dewatered ($gal/ft^2/min$)	Cake Solids Content (percent)
Shannopin	None	20/15/10	Hyflo Super-Cel	5	0.294	22.9
Shannopin	Nalcolyte 673	20/15/10	Celite 501	4	0.550	11.4
Banning	None	20/15/10	Celite 501	5	0.472	17.2
Banning	Coagulant 2350	20/15/10	Hyflo Super-Cel	4	0.527	15.5
Edgell	None	20/15/10	Celite 501	4	0.223	35.1
Edgell	Hercofloc 831	20/15/10	Celite 501	5	0.353	30.0

The filtrate produced, as is typical with a precoat, was of excellent clarity. The filtrate from rotary precoat vacuum filtration could be discharged into a stream without any further treatment.

Pressure Filtration

General

Pressure filtration is a process in which the slurry to be dewatered is forced into the filter press under high pressure and a porous media in the press retains the solids while allowing the liquid to pass. A filter press generally consists of a series of chambers with plates on either side. The plates are covered with a suitable filtering media and have a drainage surface pattern of grooves, cones, diamonds, or other forms through which the filtrate may pass to the discharge line. The chambers are formed by either recessing the plates or placing distance spacer frames between them. The plates and frames are closed together with sufficient pressure, normally applied either hydraulically or by powered screws, to seal the faces and prevent leakage.

After the slurry enters the chamber under pressure and the filtrate passes through the filter media and is discharged, the press is opened and the cake removed.

Test Apparatus - Description

The bench scale experiments were performed with a D. R. Sperry and Company stainless steel 3-5/8 inch laboratory filter press, as shown in Figure 18. The press was designed to duplicate the operation of one-half of a filter press cell. The filter media was placed over the plate at the bottom of the unit, the sludge poured into the chamber above the filter media, the upper plate with gasket added and the screw tightened to prevent leakage. The pressure for filtration was supplied by air from a tank of compressed air.

Procedure

Initial tests performed by D. R. Sperry and Company and later confirmed by this work showed that a precoat would be necessary for the efficient operation of the filter press when dewatering acid mine drainage sludge. The use of a precoat eliminates the need to recycle part of the filtrate, improves filtrate clarity and eases cake discharge.

Pressure filtration tests were initiated by precoating the pressure filter. This operation was done by adding enough of a 2 percent precoat

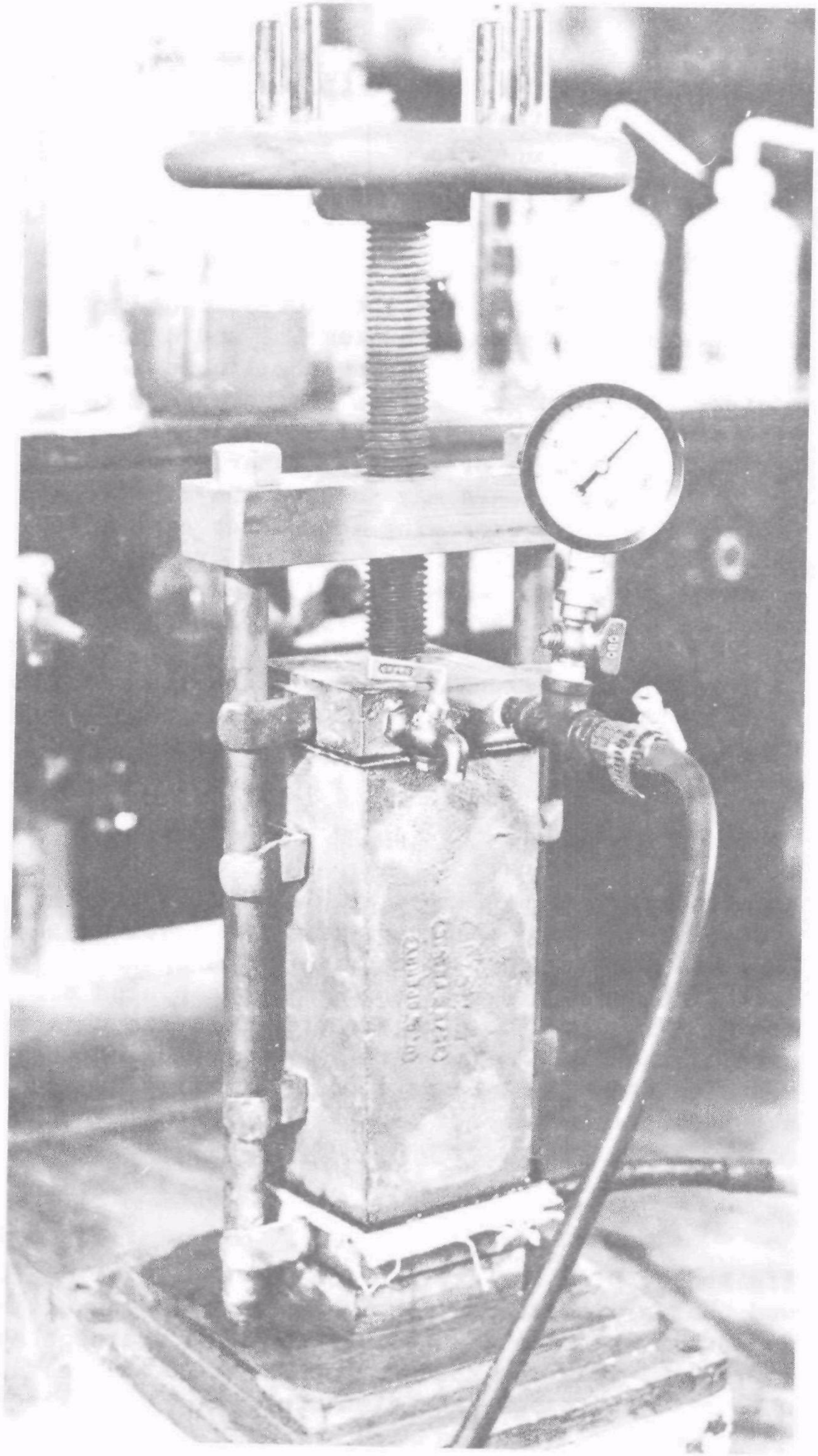


Figure 18 - PRESSURE FILTRATION APPARATUS

slurry (by weight) to the filter to form a cake approximately 1/8 inch thick as this size precoat was found to be the thinnest precoat which would not be eroded when the sludge was added to the pressure filter chamber. Pressure was applied to the precoat slurry and the slurry filtered to dryness. Next a measured amount of sludge was carefully added to the pressure filter so as not to disturb the precoat, pressure was applied, the time recorded and all the filtrate produced was collected in graduated cylinders. The amount of filtrate collected in the graduated cylinders was recorded at intervals.

As the tests conducted were of a fairly long duration, it was necessary to periodically release the pressure, open the unit and add a measured amount of additional sludge. The fresh sludge was always added before the level of the sludge in the filter dropped below the top of the sludge cake so as not to disrupt the continuity of the test.

A test was near completion when either one of two conditions were met: (1) the pressure filter chamber filled with sludge, (2) a drop or leveling off of filtrate recovery rate occurred.

When either of the conditions were met the pressure was continued until the filter cake cracked. After the cake cracked, air was allowed to flow through the cake in order to increase cake dryness. At the end of each test the cake was carefully removed from the chamber and the precoat scraped off. The cake was then measured for thickness and weighed.

Pressure variation, the use of flocculants and the length of the filter run were the variables studied in these pressure filtration tests. It was learned from talks with Calvin Mohr, a representative of D. R. Sperry and Company and an expert in the area of pressure filtration, that filtration rates could drop off at high pressures.[12] This was due either to blinding of the precoat and/or to greater cake compression which caused a decrease in cake permeability and porosity.

In order to select the optimum operating conditions, various pressures were examined and three pressures were chosen. The lower pressure, 60 psig, was the lowest pressure that could produce sufficiently high filtration rates to dewater the large volumes of sludge created by acid mine drainage treatment. The highest pressure, 100 psig, was the highest pressure at which the laboratory filter press could be safely operated. The third pressure, 80 psig, represented an intermediate pressure between the high and low pressures.

The second important variable investigated was the use of a flocculant to increase filtration rates. The most efficient flocculant at its optimum concentration for each sludge, as determined by Buchner funnel tests, was examined. Each sludge was investigated with and without flocculants and the results compared.

During each test, the amounts of sludge added to the filter and the amounts of filtrate recovered with increasing time were recorded. As the cake increased in thickness, the resistance of the cake to the passage of filtrate increased, eventually reaching a point at which the time required to remove the cake would be more than offset by the increased filtration rates obtained after its removal. Establishment of this point gave the optimum length (time) of a filter run. Having established the time of a filter run and the volume of sludge dewatered, it was possible to use this data to determine the size of a filter press necessary to dewater a given volume of sludge in a given length of time. At the end of each test, the cake was weighed and measured. From this data, the volume of cake produced by a given volume of sludge, under the conditions of the test in question, and the density (weight per unit volume) of that cake were calculated. The volume of cake produced must be known for the proper selection of a pressure unit in order to allow sufficiently large chambers to contain the cake. The density of the cake was used to calculate the weight which the unit and its supporting foundation must bear. The percent solids of the cake was determined in order to evaluate the dewatering efficiency of the variables under study and to allow for a better comparison of the various dewatering methods.

Results

Pressure filtration was found to be a feasible dewatering system for all four sludges. The results of tests performed with and without flocculants added to the four representative sludges at a low pressure (5 psig) are shown in Table 18. This table shows that the addition of flocculants gave a large increase in filtrate recovered in one minute with the final percent solids of the cake being slightly lowered.

However, after talks with and testing by D. R. Sperry and Company it was decided that sufficiently high rates could be obtained without the use of flocculants. Also pressures lower than 60 psig should not be considered since the main advantage of pressure filtration is high filtration rates per unit of filter surface area which requires high pressures. Low pressures were inefficient from a cost standpoint. Preliminary tests by D. R. Sperry and Company indicated that a precoat was necessary due to the fine particle content of the sludge. Recommendations were made that Johns-Manville Hyflo Super-Cel precoat be used along with Sperry No. 3 Cotton Twill Filter Cloth with Sperry No. 11 Filter Paper placed over it. Therefore these filter media were used in all pressure filtration tests.

Results of long duration tests are shown in Figures 19 through 22

Table 18

Pressure Filtration Tests at 5.0 psig

Sludge	Flocculant	Filtrate In One Min. (mls)	Increase In Filtrate Recovered With Floc. (percent)	Final Solids (percent)
Norton	None	72		26.3
Norton	Decolyte 940	104	44.4	25.8
Edgell	None	29		17.2
Edgell	Hercofloc 831	32	10.3	16.0
Banning	None	71		10.8
Banning	Coagulant 2350	145	104.2	8.8
Shannopin	None	20		17.7
Shannopin	Nalcolyte 673	27	35.0	16.4

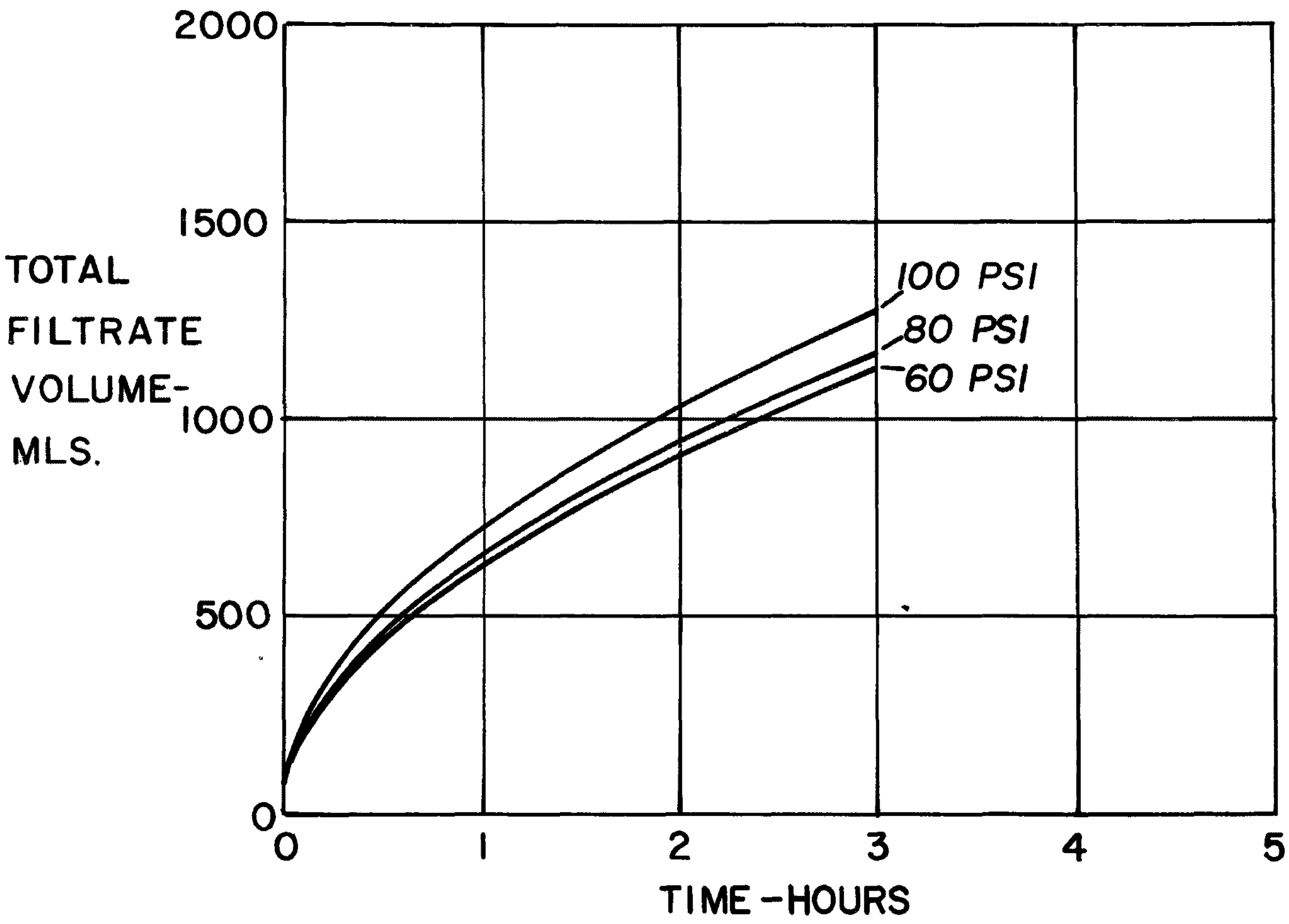


Figure 19 - PRESSURE FILTRATION TESTS OF SLUDGE FROM SHANNOPIN TREATMENT PLANT

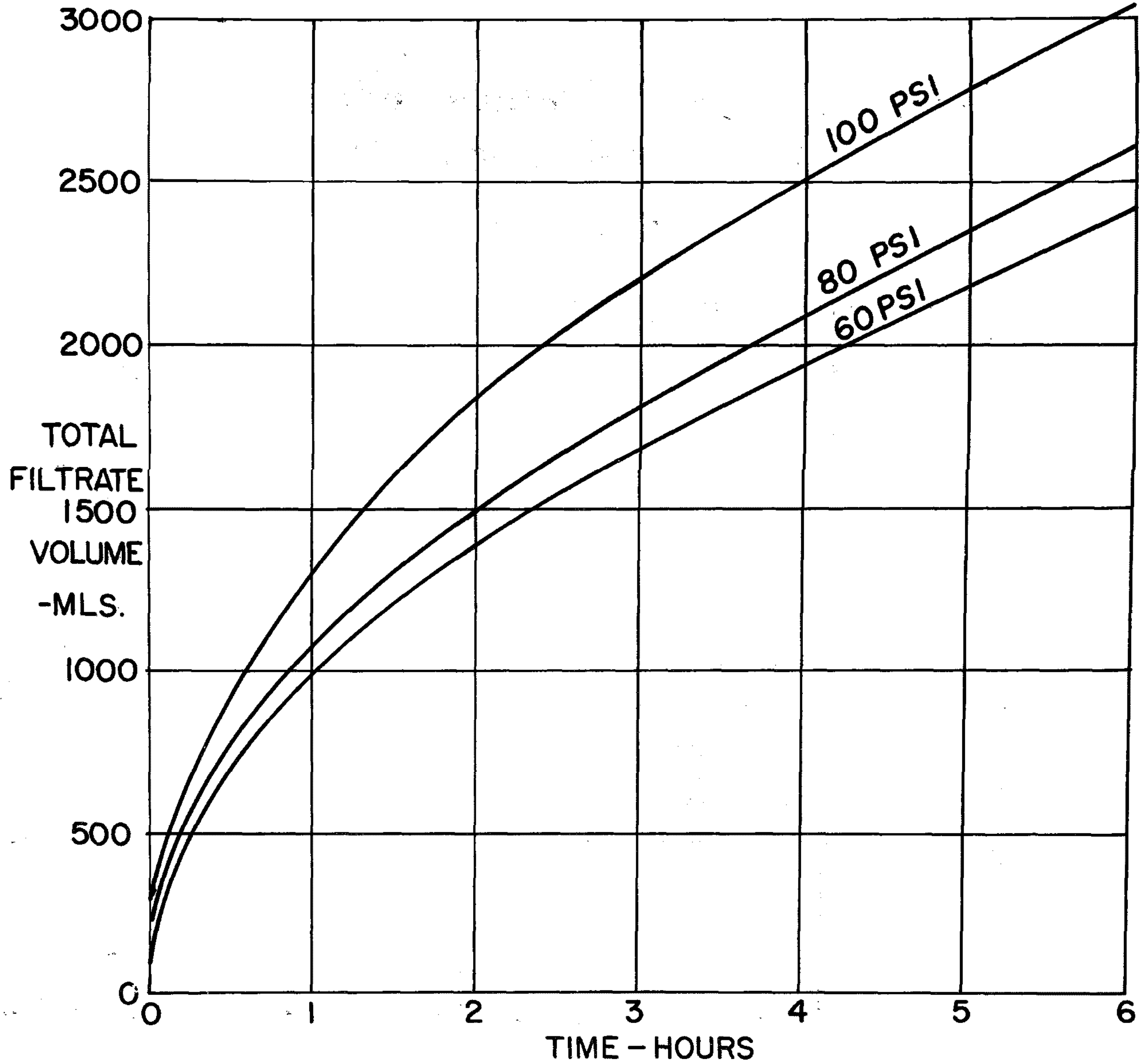


Figure 20 – PRESSURE FILTRATION TESTS OF SLUDGE FROM BANNING TREATMENT PLANT

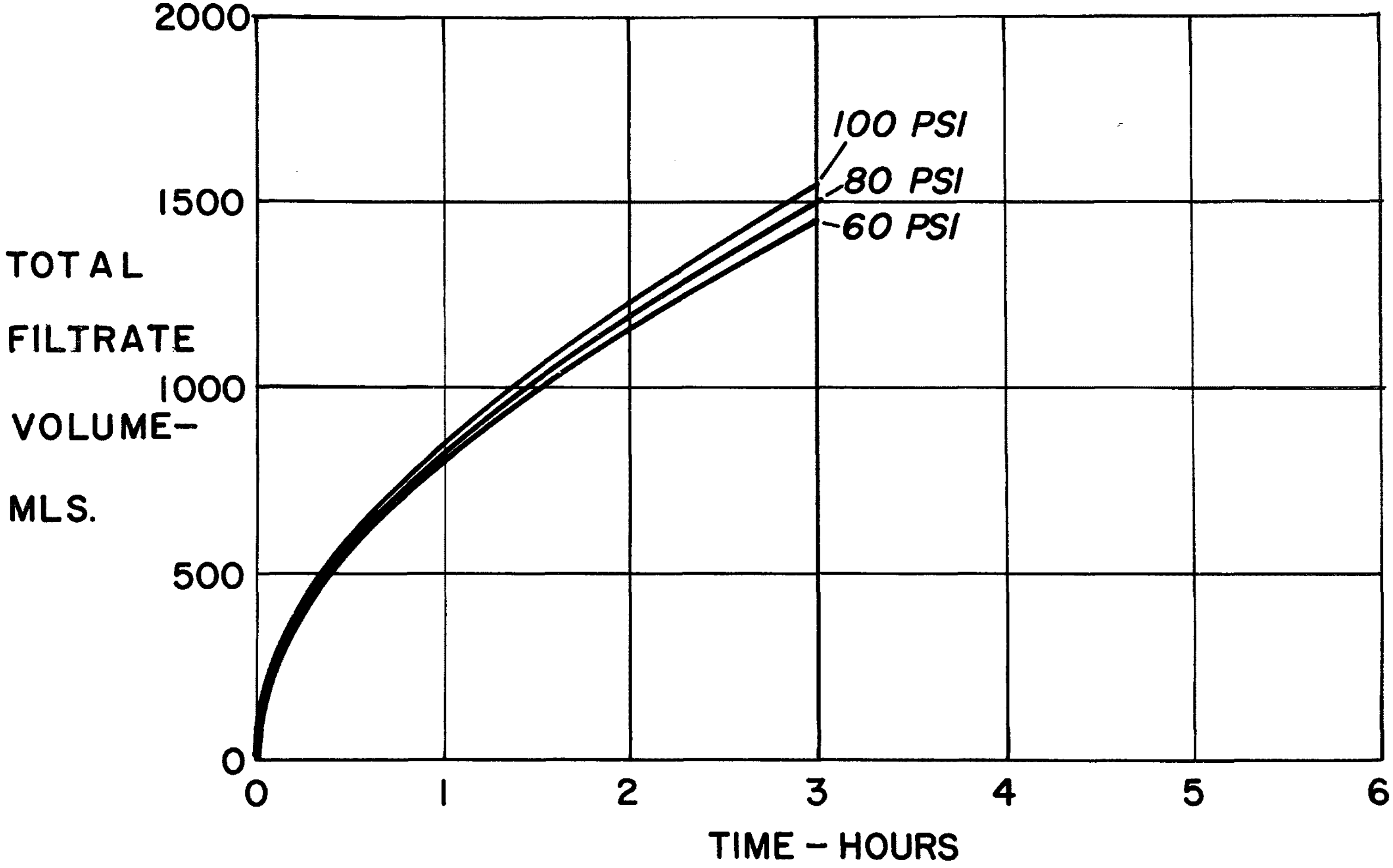


Figure 21 – PRESSURE FILTRATION TEST OF SLUDGE FROM NORTON TREATMENT PLANT

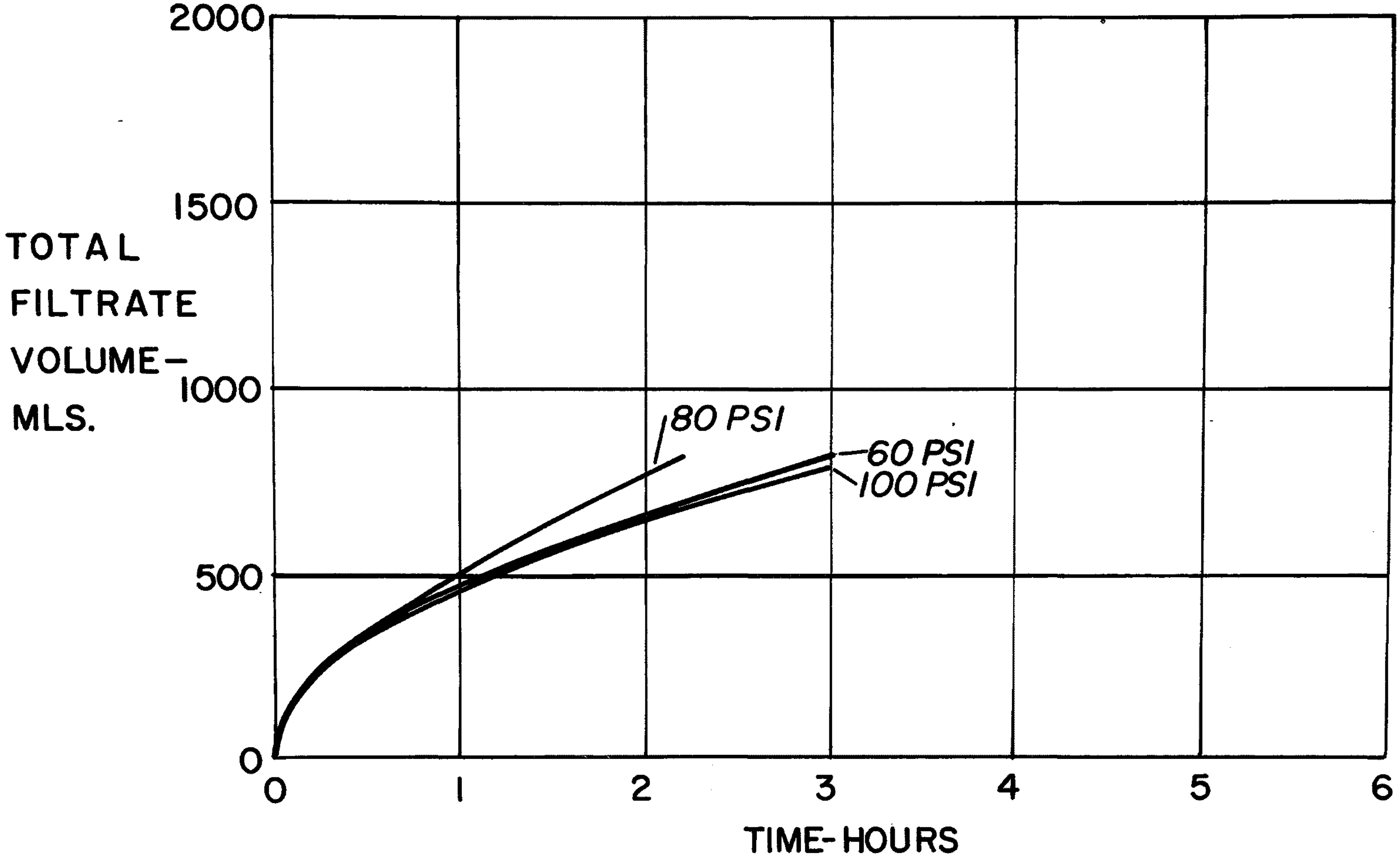


Figure 22- PRESSURE FILTRATION TESTS OF SLUDGE FROM EDGELL TREATMENT PLANT

and in Table 19. All four sludges were tested at 60, 80 and 100 psig.

Since Shannopin sludge showed nearly the same filtration rates at all three pressures, 60 psig was selected as the best operating pressure.

For Banning sludge 100 psig was selected as the best operating pressure due to the large increase in filtration rates at this pressure as compared to filtration rates at 60 and 80 psig. 100 psig was the only pressure at which cake data was taken for Banning sludge. The cake data is presented in Table 19.

Norton sludge produced a very thick cake and had nearly the same filtration rates at all three pressures. Therefore 60 psig was selected as the most economical operating pressure. This is the only pressure reported for Norton in Table 19.

With Edgell sludge the filtration rate dropped off at 100 psig due to either blinding of the precoat and/or to greater cake compaction which caused a decrease in cake permeability and porosity. The best operating pressure for Edgell sludge appeared to be 60 psig, as only a relatively small increase in filtrate recovery was realized by increasing the pressure to 80 psig.

In all tests performed a clear filtrate was recovered.

Porous Bed Filtration

General

Drying beds utilizing sand, coal, or other filtering media have been used to successfully dewater sewage and industrial sludges. Drying beds are generally constructed to hold a graded (a gradual change in particle size in a vertical direction) filtering material. A bed of coarse gravel is frequently laid over an under drain and then followed with a finer material. The finest material (placed on top of the bed) must be of sufficiently small particle size that the voids between particles are smaller than the particles of the sludge being filtered. When sludge is introduced onto the surface of the sludge drying bed, drainage immediately takes place as the filtrate percolates through the filter media and the solid particles are trapped on the surface. Following the initial drainage stage, evaporation then takes place. As the sludge becomes dryer, a point of dryness is reached when the sludge can be lifted from the surface of the drying bed. Depending upon the type of sludge being dewatered, conditioning agents can be added that accelerate the drainage cycle. To prevent rain from entering the drying bed, a covering can be placed over it.

Table 19

Pressure Filtration Cake Data

Sludge Used	Pressure (lbs/sq. in.)	Thickness of Final Cake (inches)	Filtration Time To Produce Cake (minutes)	Final Solids Of Cake (percent)
Norton	60	7.5000	179	20.8
Edgell	60	0.5000	198	26.2
Edgell	80	0.3750	138	26.0
Edgell	100	0.5625	175	26.0
Banning	100	0.8750	219	11.8
Shannopin	60	2.3750	170	8.7
Shannopin	80	2.7500	180	12.0
Shannopin	100	2.8750	200	10.9

All sludges were air blown for 5 minutes after break.

Test Apparatus - Description

The test apparatus used for the drying bed experiments was constructed of clear plastic and is shown in Figure 23. Essentially this apparatus was a tapered tank with a valve fitted at one end.

Preliminary tests were performed to find the smallest filter particle size that would completely stop the mine drainage sludge from penetrating the surface. It was found that 40 x 60 mesh wet screened sand or coal could serve as the top layer of the filter bed.

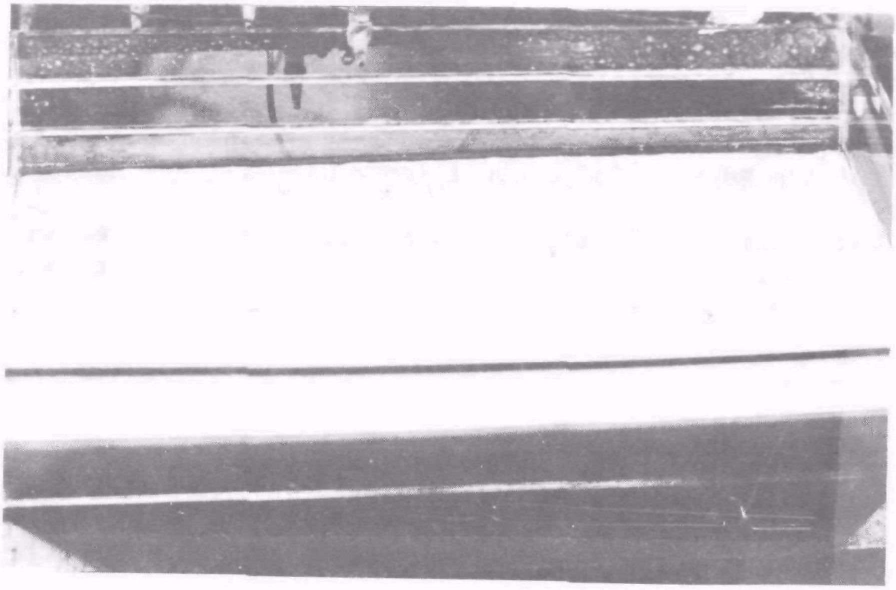
Coal was the first filter media examined. A wet screened 1.5 inch thick bed of 1 x 1/4 inch bone coal was laid down followed by 1.5 inches of 1/4 inch thick bone coal. The lower layer of coal, placed atop the bone coal, was 10 x 20 mesh, the second 20 x 40 mesh and the top 40 x 60 mesh. This layering of progressively smaller size fractions formed an approximately 6 inch thick filter bed grading from coarse at the bottom to fine at the top.

Following the filtration tests using coal as the filter media, high quartz silica sand was examined. The base of the vessel was covered by 1.5 inches of screened 1 x 1/4 inch gravel followed by a 1.5 inch layer of 1/4 inch x 10 mesh gravel. Three one inch layers of sand were then deposited on the gravel base. The lower layer was 10 x 20 mesh, the middle layer was 20 x 40 mesh and the top layer was 40 x 60 mesh. All mesh sizes given are U. S. Standard sieve series. The total filter thickness was 6 inches which duplicated the thickness and grading of the coal filter. A glass tube was inserted into the filter media at one end of the vessel to vent the filter system and allow the filtrate collected at the bottom of the filter to drain freely. The effective filter area in the test vessel was approximately 4 square feet.

Procedure

A test was initiated by first filling the vessel with tap water so as to moisten and protect the bed from erosion. The sludge was then allowed to flow from a large container by gravity onto a splash pan located in the center of the filter bed. The splash pan was introduced as part of the apparatus to minimize bed surface disturbances as the sludge was introduced. As soon as sufficient sludge was on the bed to prevent erosion the splash pan was removed.

Two tests were conducted on the coal bed filter. In each test, a 40,000 ml sample of Edgell sludge containing 6 percent solids was introduced into the vessel. One test was conducted with sludge that had been conditioned with 111 ppm Hercofloc (shown to be the best flocculant for this sludge from the conditioning studies) while the



(a)

(b)

Figure 23- POROUS BED FILTRATION APPARATUS

(a) TOP & FRONT VIEW (b) SIDE VIEW

second test was conducted with unconditioned sludge.

A series of three tests was conducted using sand as a filter media. The first test was performed with a 40,000 ml sample of 6 percent solids Edgell sludge. A second test was conducted using a 2.3 percent solids Edgell sludge and a third test was performed using 2.7 percent solids Edgell sludge conditioned with 111 ppm Hercofloc.

During each test the clarity of the filtrate was observed and a percent solids determination of the sludge cake was made at various times following the introduction of the sludge into the drying vessel.

Results

From the preliminary tests conducted, porous bed filtration appeared to be a technically feasible method of dewatering coal mine drainage sludge.

Data collected from the sand bed and coal bed drying tests is presented in Table 20. Using an arbitrary figure of 20 percent solids of the dried cake as a criteria for liftable conditions, it appears from Figure 24 that there is a linear relationship between solids loading (weight of solids per unit of surface area) and drying time.

The drying tests conducted using flocculant conditioned sludges indicated that conditioning did not significantly improve sludge dewaterability in this case. The flocculants used were selected on the basis of increased filtration rates and would be expected to speed up the filtration rate (or draining) of the sludge. However, the filtration phase of porous bed drying is short compared to the evaporation phase and even a relatively large decrease in the time required for the filtration phase has only a slight effect on total drying time.

It was noted that shortly after the drainage or filtration phase was completed, cracking was observed in the cake. This was due primarily to shrinkage of the sludge cake during drying and is similar to "mud cracks".

From the point of view of increasing the rate of drying, cake cracking appears to hold two advantages. First, as the number of cracks increase, the total surface area of the sludge cake exposed to atmospheric conditions increases. Secondly, if drying beds were used out of doors, rain water would be able, at least in part, to move through the cake surface cracks and filter bed rather than sit on the surface of the cake. There does not appear to be any advantage between either coal or sand as a filter media. No significant difference in drying time

Table 20

Data From Drying Bed Tests

	Test No.	Solids Loading (lb/sq. ft.)	Temperature (°F) ± 5	Drying Time** (Days)	Volume of Sludge Dewatered (ml)
Coal Bed	1	1.190*	75	26	40,000
Coal Bed	2	1.026	75	25	35,000
Sand Bed	3	1.010	75	24	40,000
Sand Bed	4	0.403	75	13	40,000
Sand Bed	5	0.0379*	75	12	40,000

* Treated with 111 ppm Hercofloc 831 flocculant.

** Time to achieve 20 percent solids content in cake.

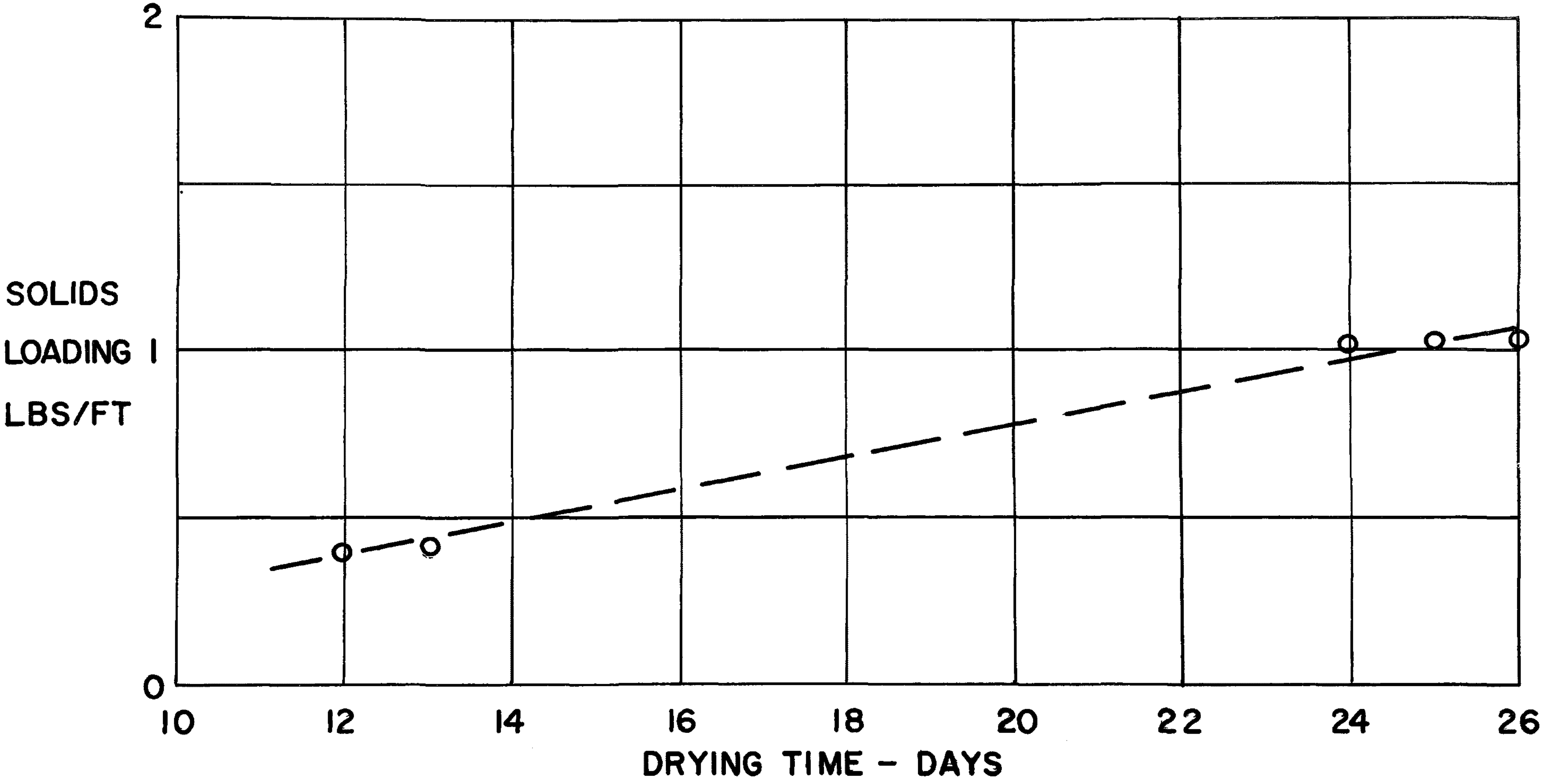


Figure 24 - RELATIONSHIP OF DRYING TIME AND SOLIDS LOADING FOR SLUDGE FROM EDGELL TREATMENT PLANT

or filtrate quality was observed.

The quality of the filtrate from both the coal filter and sand filter was quite high as indicated by an absence of non-filterable solids in the filtrate.

Thermal Spray Drying

General

Thermal spray drying is a technique that uses contact with hot gases to remove moisture from solids. It involves the following operations: (1) atomizing the sludge; (2) combining hot gases with atomized sludge droplets; (3) collecting and separating the dried product and the air.

Since spray drying operates essentially by evaporating the water from the sludge directly, the single most important cost variable is the amount of water to be removed from the sludge or the evaporative load. The costs of this system, for a given material, tend to vary inversely with the water content of the feed sludge. In order to minimize costs, a spray drying apparatus, especially for a low solids sludge, would most likely be preceded by a mechanical sludge thickening operation.

Test Apparatus - Description

Spray drying tests were conducted using a Bowen Conical Laboratory Spray Dryer, which is shown diagramatically in Figure 25.

Air for drying was heated by natural gas burners placed under a hood after which the air passed into the drying chamber.

The conical drying chamber was approximately 6 feet high, 30 inches in diameter at its widest point and constructed of type 316 stainless steel. The drying chamber had an open, uncluttered interior and an included angle of 40° to prevent an accumulation of the dried product which would be detrimental to efficiency by interfering with smooth passage of air and dried product from the chamber.

The cyclone collector was conical in shape and had roughly one-fourth the volume of the drying chamber. The dried product was discharged from the bottom of the collector and the air from the top.

Procedure

Each sludge was atomized through a spray nozzle at the top of the

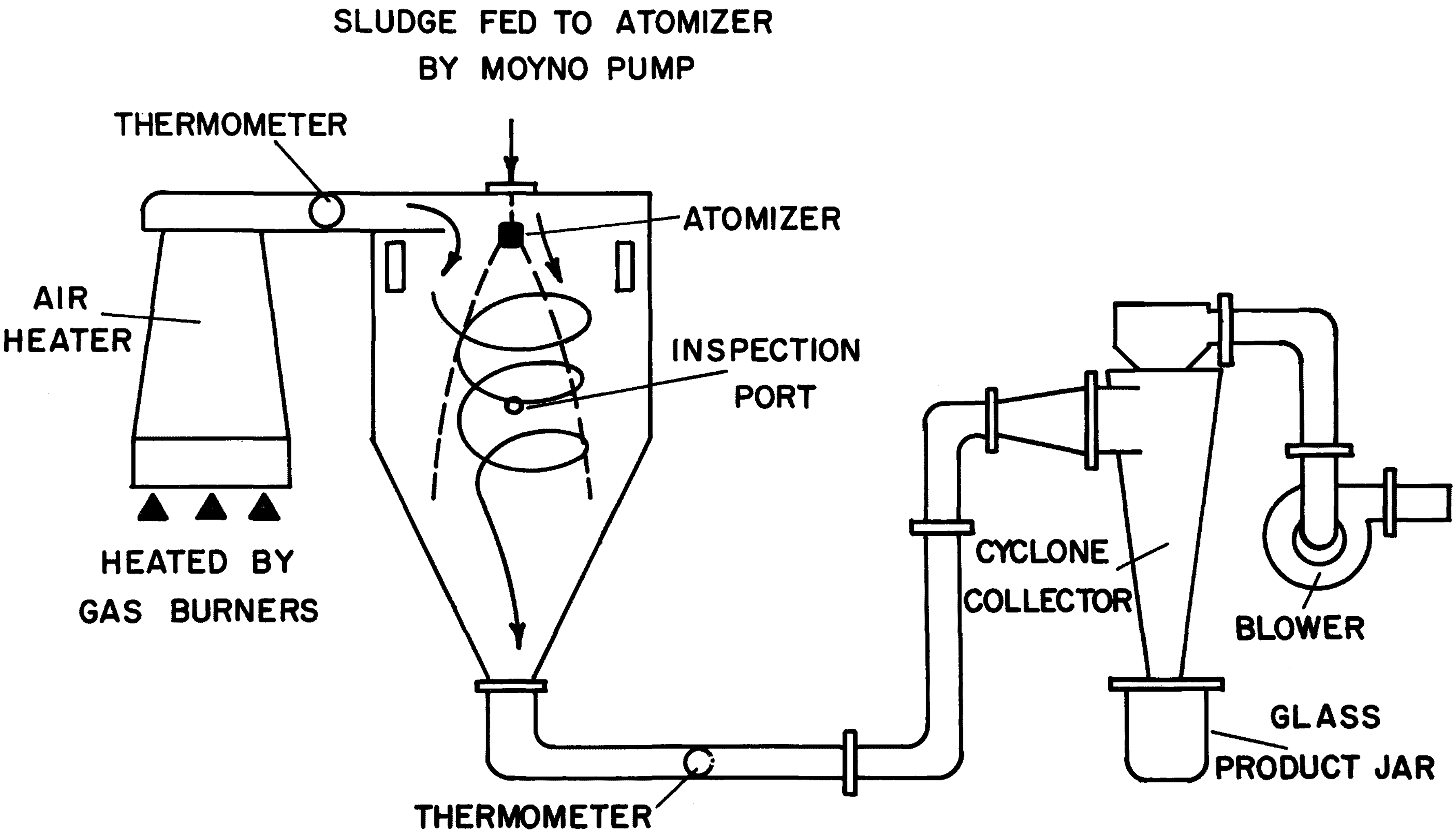


Figure 25 - SPRAY DRYING APPARATUS AFTER BOWEN ENGINEERING [13]

conical chamber. Sludge was delivered to the atomizer by a Moyno pump. Several types of atomizers are available; however, the standard model with six air holes surrounding a central hole through which the sludge was pumped, proved adequate for mine drainage sludge. The seven hole nozzle was covered with a rounded cap with a single hole in the center through which the atomized sludge was sprayed into the drying chamber.

Hot air was drawn into the conical drying chamber near the atomizing nozzle by a blower located on the hot air exhaust portion of the apparatus. This same blower pulls the air (unheated) used to atomize the sludge through the spray nozzle. The hot air inlet is so positioned as to cause the air to travel through the drying chamber in a spiral configuration. This spiraling effect increases retention time of the hot air and sludge in the chamber and promotes good mixing. Hot air which is now thoroughly mixed with the finely divided dry sludge particles passes out of the bottom of the chamber, through duct work, and into a cyclone collector where the sludge particles are removed. The hot air then passes through the blower and is exhausted. If it should prove necessary, a wet scrubber could be installed at this point to remove any remaining sludge particles or perhaps a second cyclone collector could be utilized.

In the laboratory tests performed at Bowen Engineering, Inc., the temperature of the heated air entering the conical chamber was held at 1000°F (the maximum possible with the equipment used). The temperature of the air leaving the chamber was modified by varying the flow rate of the sludges. Laboratory tests established the lowest exit temperature (or highest flow rate) which would allow the sludge to dry sufficiently so that it would not stick to the chamber walls or the duct work and clog the equipment. This lowest exit temperature represented the most efficient use of heat possible in thermal spray drying and therefore the most economical operating conditions. Data collected on the spray drying tests is summarized in Table 21.

The four sludges were tested at a solids content different from that employed in the other dewatering systems (Table 21). The solids contents used in the spray drying tests were chosen by Bowen Engineering, Inc. to allow them to better evaluate the spray drying characteristics of the sludges.

Results

All four sludges could be reduced to a very fine powder of light brown to reddish brown color with a solids content of about 90 percent. The dewatered solids content was higher than that produced by any other method under consideration. Whether or not spray drying would be an acceptable method of dewatering sludge is largely a matter of economics.

Table 21

Test Data From Spray Drying

Run No.	1	2	3	4
FEED CONDITIONS:				
Mine Drainage Sludge	Norton	Shannopin	Banning	Edgell
Wt. % Solids - Typical	8.0	2.1	0.5	2.7
Wt. % Solids Used in Test	8.5	11.0	9.5	1.0
Spec. Gravity	1.05	1.08	1.08	1.00
Temperature °F	70	70	70	70
Feed Rate Mls/Min	560-600	590-630	900-961	650-700
Total Feed, Mls	1,800	4,100	4,050	7,400
OPERATING CONDITIONS:				
Inlet Temp. °F	995	970-985	990	990-1,000
Outlet Temp. °F	340	325	265-280	305-315
Type Heat	Direct Gas Used For All			
Atomizer Type	Two Fluid Nozzle Used For All			
Atomizing Force, Air Press, PSIG	95	100	95	95

Table 21 (Continued)

Run No.	1	2	3	4
OPERATING CONDITIONS: (Continued)				
Chamber Conditions	Light Static Accumula- tion	Damp Spot Uppercone	Wet Spot on Cone and Wall	Wet Wall and Cone
MATERIAL BALANCE:				
Cyclone Collector, Gms.	105	260	145	30
Chamber Wall, Gms.	25	100	280	nil
Total Collected, Gms.	130	360	425	30
Total Solids Fed, Gms.	160.5	486	415	74
% Recovery, Wet Basis	81.0	74.1	100+	40.6

Centrifugation

General

Centrifugation is a method of separating materials of different densities by the use of centrifugal force. There are many types of commercial centrifuges but all consist of a feed system which delivers the material to be dewatered, a revolving basket or bowl to collect the solid particles, a discharge line and a system for removing the built up solids.

Material to be dewatered is fed into the bowl or basket and during the operation solids are forced away from the axis of rotation of the bowl and deposited on the wall, while the centrate fills the bowl. When the bowl is filled to capacity, the effluent is discharged, generally over the sides of the bowl, and exits through a discharge line.

The solid material is then removed by an automatic scraper blade or some other device designed to remove the material.

A solid bowl centrifuge of the type studied in this work has two methods of solids removal. The first method, skimming, is performed whenever the clarity of the effluent drops below an acceptable level. Skimming removes the softer, less compact sludge solids. The second method of solids removal involves scraping the harder, more compact sludge solids collected along the circumference of the bowl. These solids are removed when they have accumulated to the point that they have reduced the capacity of the bowl and made the time between skimming operations prohibitively short.

Test Apparatus - Description

The bench scale experiments were performed with a 14" diameter basket Fletcher Mark III centrifuge of the solid bowl type which operates at variable speeds and is electrically powered. The machine was rented from Sharples-Stokes Division of Pennwalt Corporation and is shown in Figure 26. This centrifuge was designed to duplicate a large industrial type centrifuge.

A gravitational feed system was used to supply sludge to the centrifuge. The sludge was siphoned from an elevated 55 gallon drum through a 1/2 inch pipe equipped with a globe type valve into a secondary 20 gallon cylindrical plastic container. The sludge was discharged from the secondary container into the centrifuge. Various flow rates to the centrifuge were maintained by the use of a predetermined piezometric head within the secondary container for each flow rate. The

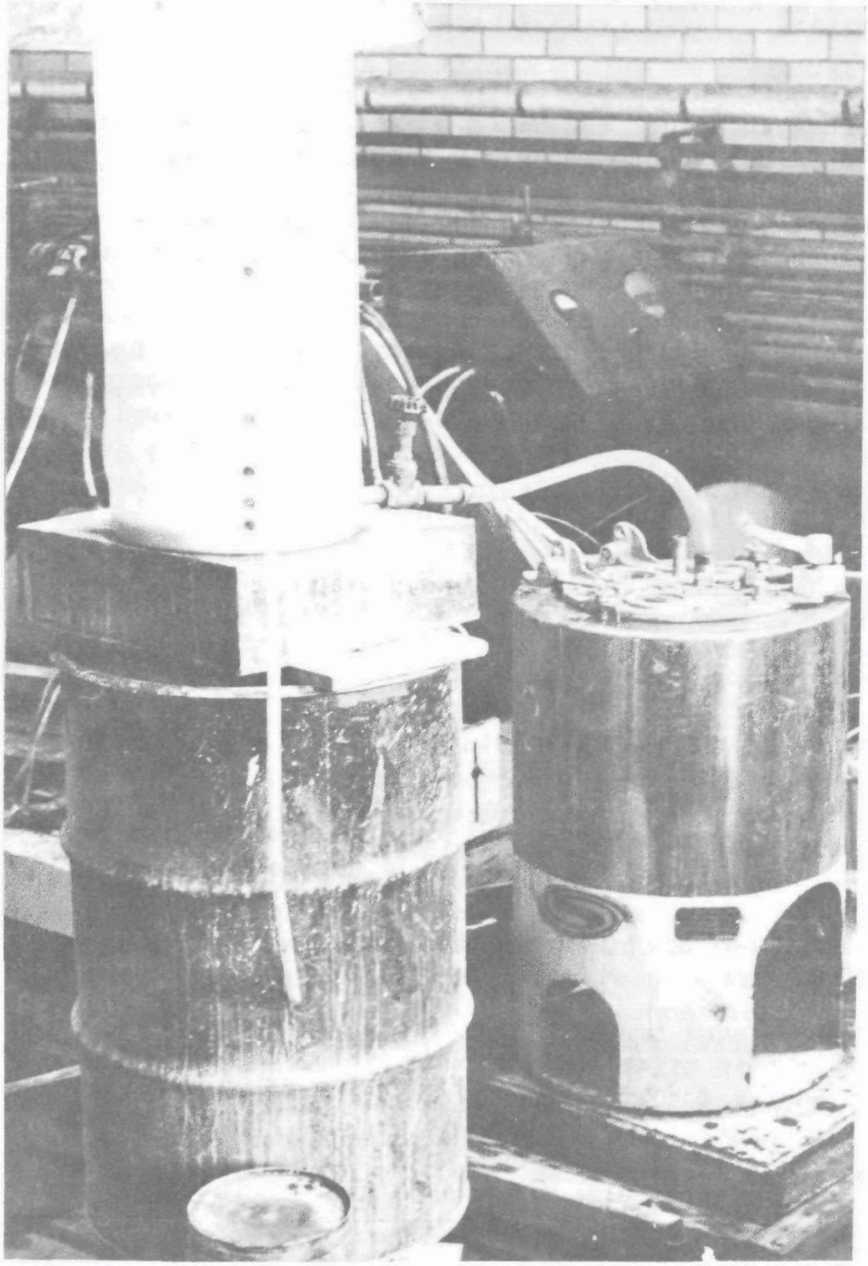


Figure 26- CENTRIFUGATION APPARATUS

sludge was kept thoroughly mixed by the use of a 1/2 horsepower electrically powered Lightnin mixer.

The solid bowl, which was cylindrical shaped, was 14 inches in diameter by 6 inches in height and had a 1 7/8 inch lip ring around the top. A truncated cone approximately 6 inches in diameter at the base and 2 inches in diameter at the top by 4 inches in height was bolted to three mounts at the bottom of the bowl. These mounts were approximately 1/4 inch in height and one inch square; therefore three slots 1/4 inch in height were situated under the conical section. Sludge entered the centrifuge through a 1 inch diameter pipe into the top of the conical section and was dispersed uniformly under the bottom of the cone; this assures flow throughout the bowl from bottom to top. Centrate flowed over the lip ring on the top of the bowl when the bowl's capacity of 1.5 gallons was reached.

The discharge line was a 2 inch hose running from the bottom of the centrifuge to the drain.

The sludge skimmer pipe was 3/8 inch diameter and passed through the lid of the centrifuge with one end of the pipe (entrance) inside the bowl and pointing in the opposite direction to the bowls rotation. The exit end of the skimmer protruded vertically from the top of the centrifuge and was bent 90° to the front of the machine. A rachet arrangement was used to advance the skimmer manually from the center of the bowl towards the side while the bowl was in rotational operation. Centrifugal force in the bowl forced the thick sludge out of the pipe when the skimmer was activated.

Procedure

Sludge to be used for each test was put into a 55 gallon drum equipped with a Lightnin mixer and a siphon pipe. The mixer was started and it continued to run for the duration of the test to keep the sludge thoroughly mixed. The siphon was started and sludge was added to the secondary plastic container and the desired predetermined piezometric head for each particular run was obtained. The flow rate was checked by measuring the amount of slurry which flowed in one minute.

The variable speed centrifuge was activated and stabilized at a rotational speed of 2750 revolutions per minute. This speed was recommended by the manufacturers of the Sharples-Fletcher centrifuge, and was based on acid mine drainage sludge tests performed earlier using a limestone sludge.

At the beginning of each test the time was recorded and samples of the effluent discharged were taken at one minute intervals, starting with the initial discharge and ending with the termination of a test. As the test progressed the bowl accumulated more and more sludge which reduced its effective volume. The reduction of the bowl's effective volume reduced the retention time of the slurry which reduced the effectiveness of solids removal. The centrifuge operation was continued until the effluent discharged had approximately the same percent solids concentration as the feed going into the bowl; this is done by comparing the clarity of a bottle of sludge feed to a bottle of effluent.

Sludge feed was then stopped, but the operation of the centrifuge continued. The effluent which remained in the bowl was skimmed by advancing the skimmer pipe into the bowl until the liquid-solid interface was reached. At this point the distance from the skimmer to the outer circumference of the bowl was recorded and was a direct measure of the amount of sludge in the bowl. Most of the sludge which remained was skimmed off; however, a very compact sludge on the side of the bowl cannot be removed by the skimmer for if contact should occur between the skimmer pipe and the bowl damage would occur. The centrifuge was then stopped. The compact cake, called the bowl cake or bowl solids, which remained was measured for thickness at the top, middle and bottom of the bowl and recorded and the sludge scraped from the side of the bowl.

The most important parameters of the test were centrifugal force (speed-diameter) feed rate and characteristics of the sludge. Tests were run at flow rates of 1/2, 1, 1-1/2, 2, 2-1/2 and 3 gallons per minute for non-flocculated sludges. If the tests which were performed at lower feed rates lasted 5 minutes or less, further tests were not made at increased feed rates as, according to the manufacturers, shorter runs would be impractical. Each sludge was also tested at 1/2 gallon per minute with the correct concentration of flocculant added as determined earlier in vacuum filtration tests.

Samples of feed, effluent discharged from beginning to end of run at selected intervals, skimmer sludge and sludge deposited on the wall of the bowl were submitted for percent solids analysis. Due to the length of each run many effluent discharge samples were collected and several were discarded as they were similar in clarity.

Results

Centrifugation was found to be a feasible dewatering system for Norton, Shannopin and Banning sludges. Edgell sludge was not tested due to difficulties encountered in obtaining this sludge in sufficiently large quantities. The Edgell settling lagoon is located far from the nearest road and samples must be carried out by hand, making the collection of several hundred gallons of sludge difficult.

As would be expected all sludges show a decrease in the time the centrifuge can be operated as the feed rate of sludge increases or as the solids content of the feed sludge increases. This is due to the increase in sludge volume deposited in the bowl with time. The more rapidly the sludge accumulates, the less the effective volume of the bowl and the less effective its removal of solids.

Figure 27 shows the results of the tests on Shannopin sludge. The length of time the centrifuge can be operated before skimming decreases with increasing feed rates as shown by Curves 1 through 3. For Curves 4 and 5 the feed rate has reached the point that centrate clarity drops to an unacceptable level (below 90 percent recovery) in less than a minute. Curve 3 appears to be a good combination of feed rate and length of run and would represent the feed rate at which the greatest volume of sludge could be dewatered per unit of time.

Curve 6 shows the effect of a flocculant on Shannopin sludge. Since Curve 6 differs very little from Curve 1 (the same feed rate) it would appear that little advantage would be gained in applying this flocculant.

Figure 28 shows the results of centrifugation tests on Banning sludge. Banning sludge behaved similarly to the Shannopin sludge with Curve 3 again representing the maximum practical feed rate. The use of a flocculant in this case increased the length of time the machine could be operated (Curve 1) before skimming and would increase the amount of sludge which could be dewatered by the machine.

Figures 29 and 30 are for Norton sludge. Figure 29 shows the rate at which the length of a run decreases as the solids content of the feed sludge increases. Curve 4 represents the test made using a flocculant. Curve 4 drops before Curve 2, a test of the same flow rate and an even higher solids content, thus the flocculant decreased the capacity of the machine. Figure 30 shows the results of increasing the feed rate of Norton sludge. Curve 2 represents the maximum feed rate under these conditions. Curves 1 and 4 may be compared to determine the effect of increasing the feed solids content of Norton sludge at a flow rate of 1 gpm.

Tables 22 through 24 present the solids content of the dewatered sludge. Norton sludge dewatered by centrifugation showed the highest solids content and Banning sludge the lowest.

Summary of Results on Dewatering Attempts

It was found that it was technically feasible to dewater all four sludges by any of the six dewatering systems studied with the ex-

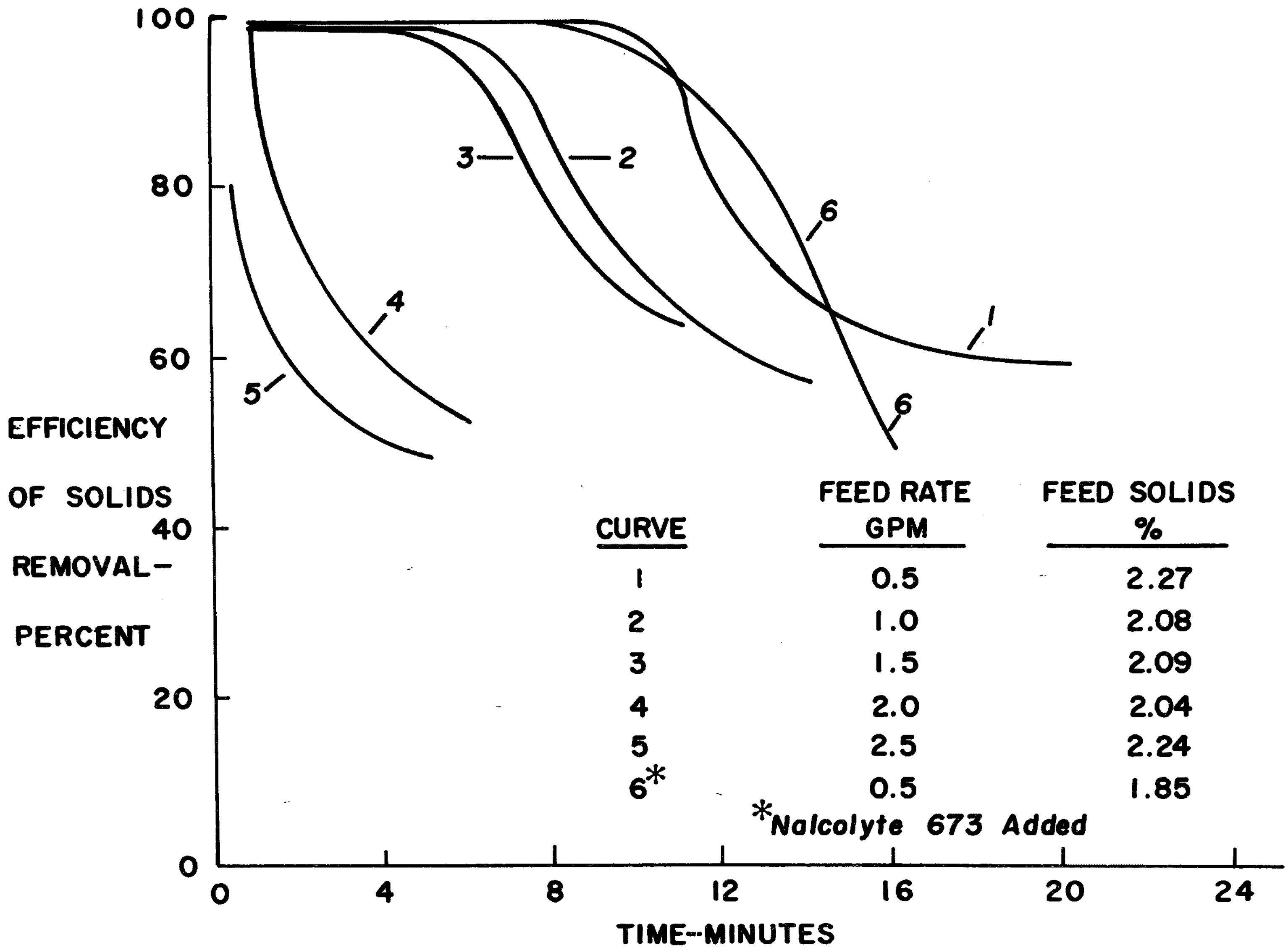


Figure 27 – CENTRIFUGATION TESTS OF SLUDGE FROM SHANNOPIN TREATMENT PLANT.

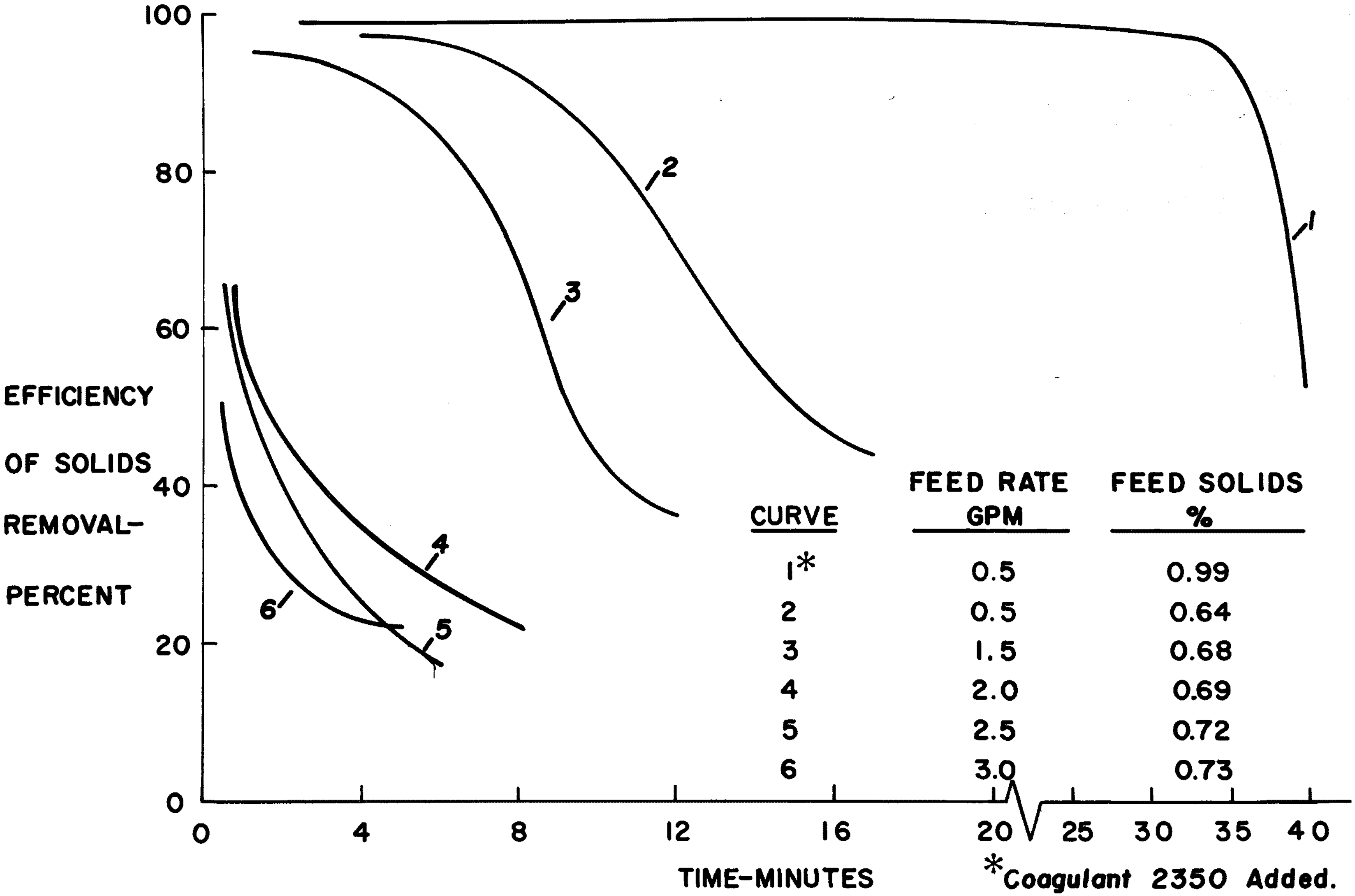


Figure 28- CENTRIFUGATION TESTS OF SLUDGE FROM BANNING TREATMENT PLANT.

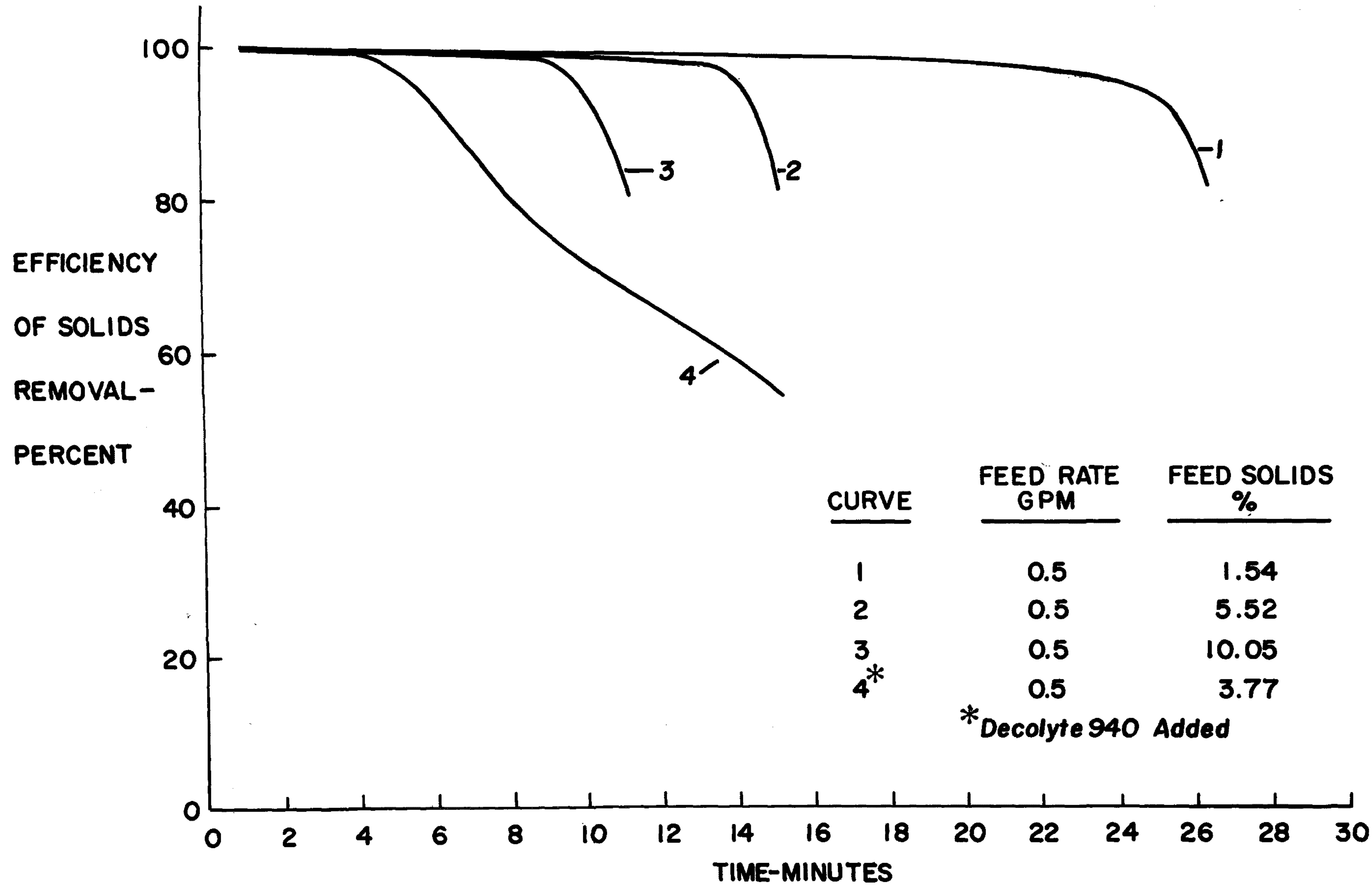


Figure 29 – CENTRIFUGATION TESTS OF SLUDGE FROM NORTON TREATMENT PLANT.

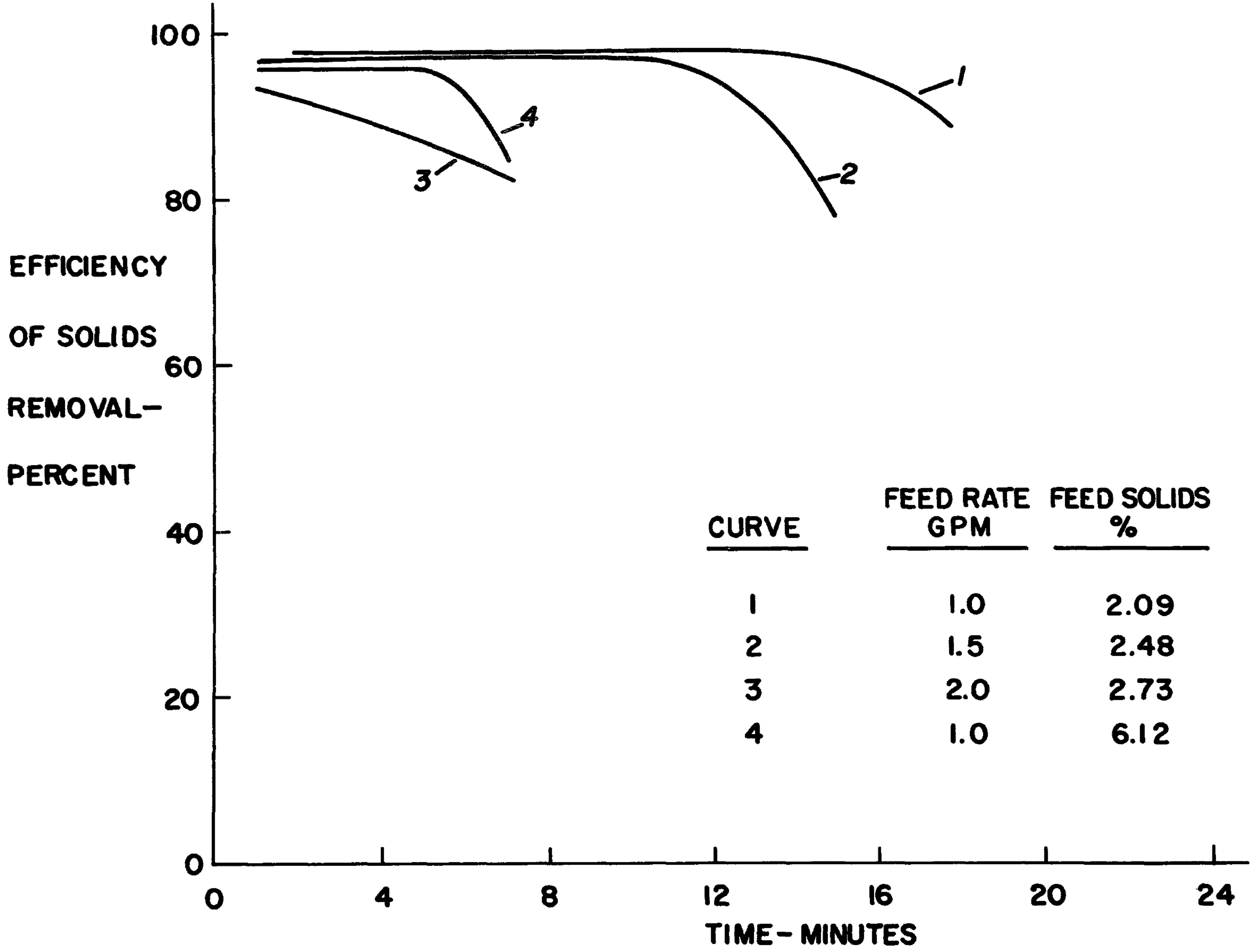


Figure 30 – CENTRIFUGATION TESTS OF SLUDGE FROM NORTON TREATMENT PLANT.

Table 22

Centrifugation Tests
Shannopin Treatment Plant Sludge

Figure Number	Curve Number	Feed Solids (percent)	Flow Rate (gpm)	Bowl Solids (percent)	Skimmer Solids (percent)
27	1	2.27	0.5	33.4	11.0
27	2	2.08	1.0	33.6	8.8
27	3	2.09	1.5	32.5	9.0
27	4	2.04	2.0	36.9	8.0
27	5	2.24	2.5	35.4	7.7
27	6*	1.85	0.5	26.6	13.8

* Nalcolyte 673 added

Table 23

Centrifugation Tests
Banning Treatment Plant Sludge

Figure Number	Curve Number	Feed Solids (percent)	Flow Rate (gpm)	Bowl Solids (percent)	Skimmer Solids (percent)
28	2	.64	0.5	8.8	5.1
28	3	.68	1.5	8.4	3.8
28	4	.69	2.0	8.7	4.6
28	5	.72	2.5	8.6	5.0
28	6	.73	3.0	8.1	5.6
28	1*	.99	0.5	11.4	9.2

* Coagulant 2350 added

Table 24

Centrifugation Tests
Norton Treatment Plant Sludge

Figure Number	Curve Number	Feed Solids (percent)	Flow Rate (gpm)	Bowl Solids (percent)	Skimmer Solids (percent)
29	1	1.54	0.5	41.3	12.9
29	2	5.52	0.5	53.7	19.3
29	3	10.05	0.5	64.1	17.5
29	4*	3.77	0.5	63.0	22.1
30	1	2.09	1.0	44.9	11.6
30	2	2.48	1.5	50.5	13.5
30	3	2.73	2.0	55.8	12.8
30	9	6.12	1.0	51.8	19.2

* Decolyte 940 added

ception that precoat rotary vacuum filtration was found not to be applicable to Norton sludge. Conventional rotary vacuum filtration was technically feasible for all sludges, but to be practical for use on Shannopin, Edgell and Banning sludges additional thickening or special equipment would be necessary.

Porous bed filtration and all systems utilizing a precoat produced a filtrate of sufficient clarity to be discharged directly into a stream. Thermal spray drying produced only water vapor which could be discharged directly into the atmosphere. Both conventional rotary vacuum filtration and centrifugation produced water of a relatively low clarity and would require recycling to the clarifier.

No system showed a significant advantage in solids content of dewatered sludge produced except for thermal spray drying. The other dewatering systems generally produced sludge cakes of from 10 to 30 percent solids while the thermal spray drying method produced dried sludge containing over 90 percent solids.

Two systems, pressure filtration and porous bed filtation were noticeably free of operational problems during bench scale testing. However, the bench scale testing of porous bed filtration did not include an evaluation of cake removal systems, a possible source of difficulty.

SECTION VII

ECONOMIC EVALUATION OF SLUDGE DEWATERING ATTEMPTS

The purpose of an economic evaluation of sludge dewatering systems is to provide a relative cost comparison of the six dewatering systems studied. To estimate highly accurate cost figures for any particular dewatering system, actual pilot plant scale tests, rather than bench scale tests, should be performed. The cost figures presented here are primarily intended to be used as guides for comparison of systems in respect to each other.

In order to make a realistic economic evaluation of the various processes, costs must be based on certain assumptions. To facilitate the best possible understanding of this analysis a complete list of the assumptions is presented:

1. Straight line depreciation over complete life of equipment.

2. Location - Morgantown, West Virginia.

3. Time - Spring of 1972

4. Operation of equipment 24 hours per day, 365 days per year with excess size to facilitate periodic shut downs.

5. Maintenance assumed to be 6 percent of total capital cost.

6. Fuel: $0.13 per gallon, #2 fuel oil, 130,800 BTU's per gallon.

7. Electricity: $0.0175 per kilowatt hour.

8. Labor Cost: $6.00 per hour, includes shift differentials, overtime premium, payroll overhead, supervision and fringe benefits.

9. Dewatering system to be constructed simultaneously with treatment facility.

10. Building: single story warehouse type structure based on $10.00 per square foot constructed.

11. Acid water from mine to be treated assumed to be:

    Shannopin 250,000 gallons/day at 4,350 ppm non-filterable solids.
    Banning 720,000 gallons/day at 2,400 ppm non-filterable solids.
    Norton, 20,000,000 gallons/day at 150 ppm non-filterable solids.
    Edgell 1,000,000 gallons/day at 2,250 ppm non-filterable solids.

12. Treatment plant clarifier from which the underflow is to be de-

watered was characterized as:

Shannopin, 180,000 gallons/day at 2.1 percent non-filterable solids
Banning, 360,000 gallons/day at 0.5 percent non-filterable solids
Norton, 50,000 gallons/day at 8.0 percent non-filterable solids
Edgell, 100,000 gallons/day at 2.7 percent non-filterable solids.

13. Retention time required for all clarifiers assumed to be one hour. A list of factors not considered in the evaluation would include:

    1. Miscellaneous supplies

    2. Laboratory costs

    3. Real estate taxes

    4. Insurance

    5. Working capital

    6. Land costs

    7. Differentials in costs of transportation and disposal due to dryness of dewatered sludge

    8. Plumbing and heating of building

Equipment specifications were determined after consultation with the respective manufacturers using extrapolated data collected during experimentation. Capital costs are totally on price quotations from the manufacturers except for porous sand bed filtration costs which are based upon a recent study by Barnard and Eckenfelder.[15]

Operational costs were based on correspondence with equipment manufacturers, material suppliers or local utilities.

As noted in the sludge dewatering section of this report, some of the dewatering systems were evaluated with a flocculant added to the sludge. The primary economic advantage obtained by the addition of a flocculant would be to increase filtration rates which would increase the efficiency of the dewatering unit. If the efficiency of a unit could be sufficiently increased, it would be possible to utilize a smaller unit thereby reducing capital costs.

Results obtained by using a flocculant were generally favorable for reducing equipment sizes. However at the concentrations determined in laboratory testing the savings in equipment would in all cases be offset in a very short time by the cost of the flocculant. Flocculant treatment of the sludges may have produced better economic results if concentrations would have been based on economic optimization rather than maximization of filtration rates.

The effluents from centrifuging and conventional vacuum filtration were felt to lack sufficient clarity for direct stream discharge and therefore would require additional treatment. The simplest and most efficient method of accomplishing this would be to recycle all of the effluent through a thickener clarifier. However, in all cases, the cost of the increased size clarifier required to handle the additional liquids did not significantly alter the final relative costs.

Conventional Vacuum Filtration

Evaluation of bench scale experiments proved that only the Norton Treatment Plant sludge (limestone sludge) possessed physical properties acceptable for dewatering by conventional vacuum filtration. This held true for sludges treated with flocculants as well as those untreated. In all cases tested, conventional vacuum filtration produced a filtrate unacceptable for discharge directly into a stream. Therefore all filtrate must be recycled through the clarifier, from which the overflow could be discharged into any nearby stream.

After correspondence with Mr. Lewis Arabia[11] of Eimco Processing Machinery Division, Envirotech Corporation, it was determined that one 6 foot diameter by 6 foot face drum filter with a belt discharge would suffice for the assumed Norton dewatering system. The cost was estimated to be $3.40 per 1,000 gallons of sludge dewatered. Adding this dewatering system to an existing neutralization system would increase total cost by $0.01 per 1,000 gallons of acid water.

Complete cost data is presented in appendices, Table 1.

Rotary Precoat Vacuum Filtration

Results of rotary precoat vacuum filtration as described in the dewatering section indicate that a cycle time of 20 seconds filtration, 15 seconds drying and 10 seconds for cutting was optimal for filtration of the three lime sludges. Norton sludge was not amenable to this dewatering system. Costs were estimated to range from $1.40 to $3.50 per 1,000 gallons of sludge dewatered. With this dewatering system added to an existing neutralization system additional costs would amount to $0.35 to $1.80 per 1,000 gallons of acid water depending on the treatment plant. Complete cost data is presented in the appendices, Tables 2 through 4.

Pressure Filtration

Laboratory experiments showed that all four sludges could be dewatered by pressure filtration. The filter media consisted of Johns-Manville Celite 501 supported horizontally by a cloth. The filter media is washed away with the dried sludge at the end of each filter cycle. The filtrate from this process was of sufficient clarity for direct discharge into a stream.

After consultation with Calvin Mohr,[12] of D. R. Sperry and Company, required press sizes were determined for the respective treatment plants. All presses were of the 48 inch EHCL type, with the number of plates and frame sizes dependent upon the particular sludge. Due to the filter sizes a plate shifter for each filter press would be required to handle the plates and frames during the cleaning process.

Two men would be required to disassemble, clean, and reassemble each system. In the systems with more than one press the operator would rotate press units cleaning each press at the end of its cycle.

Electric pumps are used to provide pressure for the filtering process.

Precoat use was based on minimum requirements without erosion, this was determined to be 0.125 of an inch in laboratory tests.

The costs were estimated to range from $1.70 to $7.30 per 1,000 gallons of sludge dewatered. On the basis of total acid water to be treated, the additional cost of dewatering would range from $0.02 to $2.40 per 1,000 gallons of acid water. Complete cost data is presented in Tables 5 through 8.

Porous Bed Filtration

Sewage sludge is generally applied to drying beds at a depth of 8 to 12 inches.[16] As no information on optimum depth for acid mine drainage sludge is available, a depth of one foot is assumed. A cubic foot of Edgell sludge at 2.7 percent solids would contain 1.685 pounds of sludge. Assuming a linear relationship between solids loading and drying time and projecting the data from Figure 24, drying time for a solids loading of 1.685 lbs/ft.^{2} is 34 days.

In a study by Barnard and Eckenfelder the capital costs of sludge drying beds were estimated at $1.15 per square foot.[15] With the assumed 100,000 gallons of sludge per day from the Edgell Treatment Plant and 34 days for a drying cycle, 454,580 square feet would be required for one complete cycle. This would require a capital cost of $522,800.00. Assuming a 10 year life for the filter bed the yearly capital cost would be $52,280.00 or $143.00 per day. At the aforestated clarifier underflow rate of 100,000 gallons per day capital costs would be reduced to $14.30 per 1,000 gallons of sludge dewatered.

Operational costs from the same study were calculated reflecting solids content. Their relationship was,

Operating cost (¢ per 1000 gallons) =

$$\frac{S \times 1.2}{Q \times 3650} \quad (0.206 + 0.94\ (1000/S^{0.5})$$

Q = flow ratio in millions of gallons per day

S = total solids in pounds per day.

With the assumed clarifier underflow at 2.7 percent solids, 22,520 pounds of solids would be produced each day. Using these rates, an operational cost of $4.80 per 1,000 gallons of sludge would be estimated.

Total capital and operational costs for porous bed filtration of the Edgell Treatment Plant sludge would be estimated at $19.10 per 1,000 gallons of clarifier underflow sludge. However, this dewatering system would increase total cost by $1.90 per 1,000 gallons of acid water.

Since the method of computation of dewatering costs used by Barnard and Eckenfelder[15] differ in some respects from that used in the other sections of this evaluation, these figures may not be directly comparable to the rest of this evaluation.

Thermal Spray Drying

Laboratory analysis of thermal spray drying was conducted by Bowen Engineering, Inc.[17] Their recommendations as to plant requirements were based on the dryer operating temperature, the quantity of moisture to be evaporated and ease of sludge moisture release.

Bowen Engineering determined the optimum operating temperatures as an inlet temperature of 1200°F and an outlet temperature of approximately 300°F. The quantity of moisture to be evaporated was extrapolated from the assumed plant volumes and percent solids. The specific gravities of the four representative sludges were not significantly larger than 1.00 (except for Norton at 1.05) so that the volume to weight ratio approximates that of water.

Cost quotations for equipment include a direct fired air heater, flame protection equipment, hot air inlet ducts, chamber outlet ducts, a cyclone collector, a scrubber, a main exhaust fan, all recording and controlling instrumentation, atomization equipment, structural support steel, all motors and start up service. Optional equipment which would be required for sludge dewatering are a feed pump, penthouse and access steel.

Most spray dryers use gas or oil direct fired air heaters; however, an indirect system can be utilized using steam. Because of the unavailability of large amounts of natural gas for commercial use in the Morgantown, W.Va. area; #2 fuel oil was chosen as the best alternative for air heating. Oil prices are based on conversation with a local marketing representative of Atlantic Richfield Company.

Power consumption is basically for the operation of the blowers and the feed pump. Additional electricity is used for control instrumentation; however, it is a small percentage of total energy requirements.

Continuous attendance during operation is not required by the systems. A full time operator is not required; however, upon advice from the manufacturer, costs were determined on the basis of one full time worker. This allows for infrequent periods when more than one worker is required. Costs were estimated to range from $15.00 to $19.00 per 1,000 gallons of sludge dewatered. Adding this dewatering system to an existing neutralization system would increase total cost by $0.05 to $11.10 per 1,000 gallons of acid water depending on the treatment plant. Complete cost data is presented in the appendices, Tables 9 through 12.

The relatively high cost of spray drying could perhaps be justified if transporting and disposing of the final dewatered sludge became a major cost factor. The spray dried sludge (approximately 90 percent solids) is much drier than sludge produced by the other dewatering methods. Since more water is removed, this sludge becomes relatively cheaper to transport and requires less area for disposal.

Centrifugation

Laboratory experiments showed that three sludges examined were acceptable for dewatering by centrifugation. Edgell Treatment Plant sludge was not tested. In all cases, the centrate produced was unacceptable for direct discharge into a stream. Therefore all centrates would have to be recycled through a thickener clarifier to obtain sufficient clarity.

After correspondence with Mr. R. A. Armstrong of Sharples-Stokes Division, Pennwalt Corporation, the required number of Sharples Sludge Pak SP-6500 centrifuges was determined for each treatment plant. The

40 horsepower drive motor is the primary consumer of electric power; however, small additional amounts are required for instrumentation and control.

Costs were estimated to range from $1.80 to $4.50 per 1,000 gallons of sludge dewatered. The increase in total cost of this dewatering system over an existing neutralization system would be from $0.01 to $1.73 per 1,000 gallons of acid water. Complete cost data is presented in the appendices, Tables 13 through 15.

Summary

From the presented cost data it can easily be observed that no one method of dewatering was absolutely advantageous to all acid treatment plant sludges. Dewatering method selection must be related to the particular treatment plant, taking into consideration variables such as acid water compositions, method of neutralization, quantity of sludge produced and final disposal of the dewatered sludge. With respect to each plant, significant cost differences occurred in both costs based on gallons of sludge and gallons of acid water. Summation of cost data for each treatment plant is presented in Tables 25 through 28.

Based on both the assumed amount of sludge and the amount of acid water, centrifugation was deemed most economical for Shannopin sludge. This sludge could be dewatered at a cost of $1.80 per 1,000 gallons of sludge or $1.30 per 1,000 gallons of acid water.

Banning Treatment Plant sludge could be dewatered at $1.40 per 1,000 gallons of sludge by both rotary precoat vacuum filtration and centrifugation. However, on the basis of acid water the cost would be $0.70 per 1,000 gallons of acid water.

Norton sludge dewatering would cost $3.30 per 1,000 gallons of sludge by conventional vacuum filtration. However, either conventional vacuum filtration or centrifugation systems could dewater the sludge at a cost of $0.01 per 1,000 gallons of acid water.

It should be noted again here that only the Norton Treatment Plant sludge possessed physical characteristics required for the conventional vacuum filtration system of dewatering. Therefore no economic evaluation was made with respect to the other three sludges using the conventional vacuum filtration system of dewatering.

Edgell Treatment Plant sludge was most economically dewatered using the rotary precoat vacuum filtration method. Based on sludge volume dewatering cost was $3.50 per 1,000 gallons of sludge. Using acid water as a criteria the cost was $0.35 per 1,000 gallons of acid water.

Variations in cost occur between analyses based on sludge and acid water because of assumptions made as to their respective solids content. Thickening results in a denser sludge that requires less dewatering to reach relative dryness.

The cost figures presented in Tables 25 through 28 are both for 1,000 gallons of sludge dewatered and as an add on cost per 1,000 gallons of acid water treated. Care must be taken in comparing the different dewatering systems on the basis of an add on cost per 1,000 gallons of acid water treated. As a result of increasing the clarifier size, or the retention time of the clarifier, the volume of sludge can be reduced with respect to the original acid water volume. In effect, the same volume of acid water produces a thicker sludge with relatively less volume. Naturally, a thicker sludge is relatively less expensive to dewater; however, the larger clarifier was not included in the cost calculations. The clarifier is normally a part of the neutralization rather than the dewatering system. Therefore in this analysis the cost of dewatering per 1,000 gallons of acid water is biased in favor of treatment systems producing relatively thicker sludges.

Table 25

COST SUMMATION FOR SHANNOPIN TREATMENT PLANT

COSTS BASED ON ASSUMPTIONS MADE IN ECONOMIC EVALUATION SECTION OF THIS REPORT

	Capital Cost	Operational Cost	Cost per 1,000 Gallons Sludge	Cost per 1,000 Gallons Acid Water*
Conventional Vacuum Filtration**				
Rotary Precoat Vacuum Filtration	$85,800.00	$428.00/day	$2.50	$1.80
Pressure Filtration	$181,200.00	$562.50/day	$3.40	$2.40
Porous Bed Filtration**				
Thermal Spray Drying	$714,800.00	$2,591.50/day	$15.50	$11.10
Centrifugation	$381,300.00	$219,30/day	$1.80	$1.30

* Cost of dewatering the sludge produced by the neutralization of 1,000 gallons of acid water.

** System not tested with this sludge.

Table 26

COST SUMMATION FOR BANNING TREATMENT PLANT

COSTS BASED ON ASSUMPTIONS MADE IN ECONOMIC EVALUATION SECTION OF THIS REPORT

	Capital Cost	Operational Cost	Cost per 1,000 Gallons Sludge	Cost per 1,000 Gallons Acid Water*
Conventional Vacuum Filtration**				
Rotary Precoat Vacuum Filtration	$85,800.00	$499.00/day	$1.40	$0.75
Pressure Filtration	$144,900.00	$563.50/day	$1.70	$0.80
Porous Bed Filtration**				
Thermal Spray Drying	$1,434,500.00	$5,005.80/day	$15.00	$7.50
Centrifugation	$762,500.00	$282.10/day	$1.40	$0.70

* Cost of dewatering the sludge produced by the neutralization of 1,000 gallons of acid water.

** System not tested with this sludge.

Table 27

COST SUMMATION FOR NORTON TREATMENT PLANT

COSTS BASED ON ASSUMPTIONS MADE IN ECONOMIC EVALUATION SECTION OF THIS REPORT

	Capital Cost	Operational Cost	Cost per 1,000 Gallons Sludge	Cost per 1,000 Gallons Acid Water*
Conventional Vacuum Filtration	$43,500.00	$160.90/day	$3.30	$0.01
Rotary Precoat Vacuum Filtration**				
Pressure Filtration	$52,000.00	$351.20/day	$7.30	$0.02
Porous Bed Filtration**				
Thermal Spray Drying	$436,600.00	$836.80/day	$19.00	$0.05
Centrifugation	$152,500.00	$181.60/day	$4.50	$0.01

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* Cost of dewatering the sludge produced by the neutralization of 1,000 gallons of acid water.

** System not tested with this sludge.

Table 28

COST SUMMATION FOR EDGELL TREATMENT PLANT

COSTS BASED ON ASSUMPTIONS MADE IN ECONOMIC EVALUATION SECTION OF THIS REPORT

	Capital Cost	Operational Cost	Cost per 1,000 Gallons Sludge	Cost per 1,000 Gallons Acid Water*
Conventional Vacuum Filtration**				
Rotary Precoat Vacuum Filtration	$76,600.00	$331.30/day	$3.50	$0.35
Pressure Filtration	$139,500.00	$568.60/day	$6.00	$0.60
Porous Bed Filtration	$522,800.00	$143.00/day	$19.10	$1.90
Thermal Spray Drying	$576,800.00	$1,573.80/day	$17.30	$1.73
Centrifugation**				

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* Cost of dewatering the sludge produced by the neutralization of 1,000 gallons of acid water.

** System not tested with this sludge.

SECTION VIII

ACKNOWLEDGMENTS

Mr. Edwin B. Wilson, now associated with Bethlehem Steel Corporation, submitted the proposal for this project and his efforts are gratefully acknowledged.

Thanks are due to Russell W. Frum who was the author of the Economic Evaluation Section of this report and Larry G. Shaffer who authored portions of the Sludge Dewatering Section of this report. Messrs. Russell W. Frum and Larry G. Shaffer also performed much of the bench scale tests during the data collection phase of this report.

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Mr. Richard B. Muter, Mr. Kenneth K. Humphreys and Mr. William F. Lawrence are acknowledged for their aid in the preparation of the final report of this project. Mr. Richard Muter also developed the acid water collection and analysis procedure detailed later in this report and assisted in the interpretation of the results of the chemical analyses. Mr. Charles R. McFadden prepared the illustrations for this report and Mr. Lionel L. Craddock prepared the photographs. The cooperation of Miss Martha E. Fekete in typing this report is gratefully acknowledged.

SECTION IX

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SECTION X

APPENDICES

Table 29

Conventional Vacuum Filtration - Norton Treatment Plant Sludge (Preliminary Prices - Spring 1972)

Capital Costs			
6 foot diameter x 6 foot drum filter with accessories	$30,000.00		
Construction and Installation: 35 percent of equipment	10,500.00		
Total Equipment Cost	$40,500.00		
Equipment Depreciation: 10 years expected life no salvage value		$4,050.00/year	$11.10/day
Building: 300 sq. ft. at $10.00/sq. ft.	3,000.00		
Building Depreciation: 30 years expected life no salvage value		$100.00/year	0.30/day
Total Capital Cost	$43,500.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$2,790.00/year	7.60/day
Electricity: 530 KWH/day at $0.0175 KWH			9.30/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$172.30/day
Assuming 50,000 gallons clarifier underflow per day			$3.30/1,000 gallons sludge de-watered
Assuming 20,000,000 gallons acid water per day			$0.01/1,000 gallons acid water

Table 30

Rotary Precoat Vacuum Filtration - Shannopin Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
Equipment: Includes one 10 foot diameter x 17 foot face vacuum filter with all motors, vacuum pump with motor, vacuum receivers, filtrate pump with motor, precoat mix tank with motor and precoat slurry pump	$59,100.00		
Construction and Installation: 35 percent of equipment	20,700.00		
Total Equipment Cost	$79,800.00		
Equipment Depreciation: 10 years expected life no salvage value		$7,980.00/year	$21.90/day
Building: 600 sq. ft. at $10.00/sq. ft.	6,000.00		
Building Depreciation: 30 years expected life no salvage value		$200.00/year	0.50/day
Total Capital Cost	$85,800.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$5,150.00/year	$14.10/day
Electricity: 2,250 KWH/day at $0.0175/KWH			39.40/day
Labor: 24 hours at $6.00/hour			$144.00/day
Precoat: Johns-Manville Hyflo Super-Cel, $68.00/ton F.O.B. California warehouse, $100.40/ton delivered Morgantown, W.Va.			$230.50/day

Table 30 (Continued)

Total Capital and Operational Cost	$450.40/day
Assuming 180,000 gallons clarifier underflow per day	$2.50/1,000 gallons sludge dewatered
Assuming 250,000 gallons acid water per day	$1.80/1,000 gallons acid water

Table 31

Rotary Precoat Vacuum Filtration - Banning Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
Equipment: Includes one 10 foot diameter x 17 foot face vacuum filter with all motors, vacuum pump with motor, vacuum receivers, filtrate pump with motor, precoat mix tank with motor and precoat slurry pump	\$59,140.00		
Construction and Installation: 35 percent of equipment	20,700.00		
Total Equipment Cost	\$79,800.00		
Equipment Depreciation: 10 years expected life no salvage value		\$7,980.00/year	\$21.90/day
Building: 600 sq. ft. at \$10.00/sq. ft.	6,000.00		
Building Depreciation: 30 years expected life no salvage value		\$200.00/year	0.50/day
Total Capital Cost	\$85,800.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		\$5,150.00/year	\$14.10/day
Electricity: 2,250 KWH/day at \$0.0175/KWH			39.40/day
Labor: 24 hours at \$6.00/hour			\$144.00/day
Precoat: Johns-Manville Celite 501, \$73.00/ton F.O.B. California warehouse, \$105.40/ton delivered Morgantown, W.Va.			\$301.50/day

Table 31 (Continued)

Total Capital and Operational Cost	$521.40/day
Assuming 360,000 gallons clarifier underflow per day	$1.40/1,000 gallons sludge dewatered
Assuming 720,000 gallons acid water per day	$0.75/1,000 acid water

Table 32

Rotary Precoat Filtration - Edgell Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
Equipment: Includes one 10 foot diameter x 14 foot face vacuum filter with all motors, vacuum pump with motor, vacuum receivers, filtrate pump with motor, precoat mix tank with motor and precoat slurry pump	$53,000.00		
Construction and Installation: 35 percent of equipment	18,600.00		
Total Equipment Cost	$71,600.00		
Equipment Depreciation: 10 years expected life no salvage value		$7,160.00/year	$19.60/day
Building: 500 sq. ft. at $10.00/sq. ft.	5,000.00		
Building Depreciation: 30 years expected life no salvage value		$166.00/year	0.50/day
Total Capital Cost	$76,600.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$4,600.00/year	$12.60/day
Electricity: 1,880 KWH/day at $0.0175/KWH			32.90/day
Labor: 24 hours at $6.00/hour			$144.00/day
Precoat: Johns-Manville Celite 501, $73.00/ton F.O.B. California warehouse, $105.40/ton delivered Morgantown, W.Va.			$141.80/day

Table 32 (Continued)

Total Capital and Operational Cost	$351.40/day
Assuming 100,000 gallons clarifier underflow per day	$3.50/1,000 gallons sludge de-watered
Assuming 1,000,000 gallons acid water per day	$0.35/1,000 gallons acid water

Table 33

Pressure Filtration - Shannopin Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

<u>Capital Costs</u>			
4-48 inch filter presses at $20,000.00	$80,000.00		
4-plate shifters at $2,400.00 each	9,600.00		
Feed pump with accessories	900.00		
Precoat equipment at $5,000.00 each	20,000.00		
Construction and Installation: 35 percent of equipment	38,700.00		
Total Equipment Cost	$149,200.00		
Equipment Depreciation: 10 years expected life no salvage value		$14,920.00/year	$40.90/day
Building: 3,200 sq. ft. at $10.00/sq. ft.	32,000.00		
Building Depreciation: 30 years expected life no salvage value		1,066.00/year	2.90/day
Total Capital Cost	$181,200.00		
<u>Operational Costs</u>			
Maintenance: 6 percent of total capital cost		$10,874.00/year	$29.80/day
Electricity: 270 KWH/day at $0.0175/KWH			4.70/day
Labor: 24 hours, 2 men at $6.00/hour each			$288.00/day
Precoat: Johns-Manville Celite 501, $73.00/ton F.O.B. California warehouse, $105.40/ton delivered Morgantown, W.Va.			$240.00/day

Table 33 (Continued)

Total Capital and Operational Cost	$606.30/day
Assuming 180,000 gallons clarifier underflow per day	$3.40/1,000 gallons sludge dewatered
Assuming 250,000 gallons acid water per day	$2.40/1,000 gallons acid water

Table 34

Pressure Filtration - Banning Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
3-48 inch filter presses at $23,000.00 each	$69,000.00		
3-plate shifters at $2,300.00 each	6,900.00		
Feed pump with accessories	900.00		
Precoat equipment at $5,000.00 each	15,000.00		
Construction and Installation: 35 percent of equipment	32,100.00		
Total Equipment Cost	$123,900.00		
Equipment Depreciation: 10 years expected life no salvage value		$12,390.00/year	$33.90/day
Building: 2,100 sq. ft. at $10.00/sq. ft.	21,000.00		
Building Depreciation: 30 years expected life no salvage value		$700.00/year	1.90/day
Total Capital Cost	$144,900.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$8,694.00/year	$23.80/day
Electricity: 270 KWH/day $0.0175/KWH			4.70/day
Labor: 24 hours, 2 men at $6.00/hour each			$288.00/day
Precoat: Johns-Manville Celite 501, $73.00/ton F.O.B. California warehouse, $105.50/ton delivered Morgantown, W.Va.			$247.00/day

Table 34 (Continued)

Total Capital and Operational Cost	\$599.30/day
Assuming 360,000 gallons clarifier underflow per day	\$1.70/1,000 gallons sludge de-watered
Assuming 720,000 gallons acid water per day	\$0.80/1,000 gallons acid water

Table 35

Pressure Filtration - Norton Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

<u>Capital Costs</u>			
1-48 inch filter press	\$21,000.00		
1-plate shifter	2,700.00		
Feed pump with accessories	900.00		
Precoat equipment	5,000.00		
Construction and Installation: 35 percent of equipment	10,400.00		
Total Equipment Cost	\$40,000.00		
Equipment Depreciation: 10 years expected life no salvage value		\$4,000.00/year	\$11.00/day
Building: 1,200 sq. ft. at \$10.00/sq. ft.	12,000.00		
Building Depreciation: 30 years expected life no salvage value		\$400.00/year	1.00/day
Total Capital Cost	\$52,000.00		
<u>Operational Costs</u>			
Maintenance: 6 percent of total capital cost		\$3,120.00/year	8.50/day
Electricity: 270 KWH/day at \$0.0175 KWH			4.70/day
Labor: 24 hours, 2 men at \$6.00/hour each			\$288.00/day
Precoat: Johns-Manville Celite 501, \$73.00/ton F.O.B. California warehouse, \$105.40/ton delivered Morgantown, W.Va.			50.00/day

Table 35 (Continued)

Total Capital and Operational Cost	\$363.20/day
Assuming 50,000 gallons clarifier underflow per day	\$7.30/1,000 gallons sludge de-watered
Assuming 20,000,000 gallons acid water per day	\$0.02/1,000 gallons acid water

Table 36

Pressure Filtration - Edgell Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
3-48 inch filter presses at $22,500.00 each	$67,500.00		
3-plate shifters at $2,200.00 each	6,600.00		
Feed pump with accessories	900.00		
Precoat equipment at $5,000.00 each	15,000.00		
Construction and Installation: 35 percent of equipment	31,500.00		
Total Equipment Cost	$121,500.00		
Equipment Depreciation: 10 years expected life no salvage value		$12,150.00/year	$33.30/day
Building: 1,800 sq. ft. at $10.00/sq. ft.	18,000.00		
Building Depreciation: 30 years expected life no salvage value		$600.00/year	1.60/day
Total Capital Cost	$139,500.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$8,370.00/year	$22.90/day
Electricity: 270 KWH/day at $0.0175/KWH			4.70/day
Labor: 24 hours, 2 men at $6.00/hour each			$288.00/day
Precoat: Johns-Manville Celite 501, $73.00/ton F.O.B. California warehouse, $105.40/ton delivered Morgantown, W.Va.			$253.00/day

Table 36 (Continued)

Total Capital and Operational Cost	\$603.50/day
Assuming 100,000 gallons clarifier underflow per day	\$6.00/1,000 gallons sludge dewatered
Assuming 1,000,000 gallons acid water per day	\$0.60/1,000 gallons acid water

Table 37

Thermal Spray Drying - Shannopin Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
1-36 foot spray drying system	$500,000.00		
Equipment Options:			
Feed Pump	7,000.00		
Access Steel	18,000.00		
Construction and Installation:			
35 percent of equipment	$181,300.00		
Total Equipment Cost	$706,300.00		
Equipment Depreciation:			
10 years expected life		$70,630.00/year	
no salvage value			$193.50/day
Penthouse (Building)	$8,500.00		
Building Depreciation:		$23.00/year	
30 years expected life			$0.10/day
no salvage value			
Total Capital Cost	$714,800.00		
Operational Costs			
Maintenance: 6 percent of		$42,888.00/year	
total capital cost			$117.50/day
Fuel: #2 fuel oil, 16,900 gallons/day at $0.13/gallon			$2,197.00/day
Electricity: 7,600 KWH/day at $0.0175/KWH			$133.00/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$2,785.10/day
Assuming 180,000 gallons clarifier underflow per day			$15.50/1,000 gallons sludge de-watered

Table 37 (Continued)

Assuming 250,000 gallons acid water per day	$11,10/1,000 gallons acid water

Table 38

Thermal Spray Drying - Banning Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
2-36 foot spray drying systems at $500,000.00 each	$1,000,000.00		
Equipment Options:			
Feed Pump	14,000.00		
Access Steel	36,000.00		
Construction and Installation:			
35 percent of equipment	$367,500.00		
Total Equipment Cost	$1,417,500.00		
Equipment Depreciation: 10 years expected life no salvage value		$141,750.00/year	$388.40/day
Penthouse (Building)	$17,000.00		
Building Depreciation: 30 years expected life no salvage value		$567.00/year	$1.60/day
Total Capital Cost	$1,434,500.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$86,070.00/year	$250.80/day
Fuel: #2 fuel oil, 34,300 gallons/day at $0.13/gallon			$4,459.00/day
Electricity: 15,200 KWH/day at $0.0175/KWH			$152.00/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$5,395.80/day
Assuming 360,000 gallons clarifier underflow per day			$15.00/1,000 gallons sludge dewatered

Table 38 (Continued)

Assuming 720,000 gallons acid water per day	\$7.50/1,000 gallons acid water

Table 39

Thermal Spray Drying - Norton Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
1-30 foot spray drying system	$300,000.00		
Equipment Options:			
Feed Pump	3,500.00		
Access Steel	14,000.00		
Construction and Installation:			
35 percent of equipment	$111,100.00		
Total Equipment Cost	$428,600.00		
Equipment Depreciation:		$42,860.00/year	
10 years expected life			$117.40/day
no salvage value			
Penthouse (Building)	$8,000.00		
Building Depreciation:		$267.00/year	
30 years expected life			$0.70/day
no salvage value			
Total Capital Cost	$436,600.00		
Operational Costs			
Maintenance: 6 percent of		$26,196.00/year	
total capital cost			$71.80/day
Fuel: #2 fuel oil, 4,300 gallons/day at $0.13/gallon			$559.00/day
Electricity: 6,200 KWH/day at $0.0175/KWH			62.00/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$954.90
Assuming 50,000 gallons clarifier underflow per day			$19.00/1,000 gallons sludge de-watered

Table 39 (Continued)

Assuming 20,000,000 gallons acid water per day	\$0.05/1,000 gallons acid water

Table 40

Thermal Spray Drying - Edgell Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
1-34 foot spray drying system	$400,000.00		
Equipment Options:			
Feed Pump	5,000.00		
Access Steel	16,000.00		
Construction and Installation:			
35 percent of equipment	$147,300.00		
Total Equipment Cost	$568,300.00		
Equipment Depreciation:		$56,830.00/year	
10 years expected life			$155.70/day
no salvage value			
Penthouse (Building)	$8,500.00		
Building Depreciation:		$283.00/year	
30 years expected life			$0.80/day
no salvage value			
Total Capital Cost	$576,800.00		
Operational Costs			
Maintenance: 6 percent of		$34,608.00/year	
total capital cost			$94.80/day
Fuel: #2 fuel oil, 9,300 gallons/day at $0.13/gallon			$1,209.00/day
Electricity: 7,200 KWH/day at $0.0175/KWH			126.00/day
Labor: 24 hours at $6.00/hour			144.00/day
Total Capital and Operational Cost			$1,730.30/day
Assuming 100,000 gallons clarifier underflow per day			$17.30/1,000 gallons sludge dewatered

Table 40 (Continued)

Assuming 1,000,000 gallons acid water per day	\$1.73/1,000 gallons acid water

Table 41

Centrifugation - Shannopin Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs

5-Sharples Sludge Pak SP-6500 centrifuge, $55,000.00 each	$275,000.00		
Construction and Installation: 35 percent of equipment	96.300.00		
Total Equipment Cost	$371,300.00		
Equipment Depreciation: 10 years expected life no salvage value		$37,200.00/year	$101.91/day
Building: 1,000 sq. ft. at $10.00/sq. ft.	$10,000.00		
Building Depreciation: 30 years expected life no salvage value		$33.00/year	$0.90/day
Total Capital Cost	$381,300.00		

Operational Costs

Maintenance: 6 percent of total capital cost		$22,878.00/year	$62.70/day
Electricity: 720 KWH/day at $0.0175/KWH			12.60/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$322.10/day
Assuming 180,000 gallons clarifier underflow per day			$1.80/1,000 gallons sludge de-watered
Assuming 250,000 gallons acid water per day			$1.30/1,000 gallons acid water

Table 42

Centrifugation - Banning Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs

10-Sharples Sludge Pak SP-6500 centrifuge, $55,000.00 each	$550,000.00		
Construction and Installation: 35 percent of equipment	192,500.00		
Total Equipment Cost	$742,500.00		
Equipment Depreciation: 10 years expected life no salvage value		$74,250.00/year	$203.40/day
Building: 2,000 sq. ft. at $10.00/sq. ft.	$20,000.00		
Building Depreciation: 30 years expected life no salvage value		$667.00/year	$1.80/day
Total Capital Cost	$762,500.00		

Operational Costs

Maintenance: 6 percent of total capital cost		$45,750.00/year	$125.50/day
Electricity: 720 KWH/day at $0.0175 KWH			12.60/day
Labor: 24 hours at $6.00/hour			$144.00/day
Total Capital and Operational Cost			$487.10/day
Assuming 360,000 gallons clarifier underflow per day			$1.40/1,000 gallons sludge de-watered
Assuming 720,000 gallons acid water per day			$0.70/1,000 gallons acid water

Table 43

Centrifugation - Norton Treatment Plant Sludge
(Preliminary Prices - Spring 1972)

Capital Costs			
2-Sharples Sludge Pak SP-6500 centrifuge, $55,000.00 each	$110,000.00		
Construction and Installation: 35 percent of equipment	38,500.00		
Total Equipment Cost	$148,500.00		
Equipment Depreciation: 10 years expected life no salvage value		$14,850.00/year	$40.70/day
Building: 400 sq. ft. at $10.00/sq. ft	$4,000.00		
Building Depreciation: 30 years expected life no salvage value		$133.00/year	$0.40/day
Total Capital CCost	$152,500.00		
Operational Costs			
Maintenance: 6 percent of total capital cost		$9,150.00/year	$25.00/day
Electricity: 720 KWH/day at $0.0175/KWH			12.60/day
Labor: 24 hours at $6.00/hr			$144.00/day
Total Capital and Operational Cost			$222.70/day
Assuming 50,000 gallons clarifier underflow per day			$4.50/1,000 gallons sludge de-watered
Assuming 20,000,000 gallons acid water per day			$0.01/1,000 gallons acid water

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SELECTED WATER RESOURCES ABSTRACTS

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1. Report No.	
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5. Report Date	
6.	
8. Performing Organization Report No.	
7. Author(s)	David J. Akers, Jr., & Edward A. Moss
10. Project No.	EPA - 14010 FJX
9. Organization	West Virginia University Morgantown, W. Va.
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15. Supplementary Notes	**Environmental Protection Agency report number, EPA-R2-73-169, February 1973.**

16. *Abstract* A study of various acid mine drainage sludge conditioning methods and dewatering systems was made. Acid mine drainage & sludge from neutralization plants were characterized. Four sludges were selected as being representative of the various types of sludges produced by the lime/limestone neutralization of acid mine drainage.

The conditioning methods studied were: freezing, use of flocculants, and use of filter aids. The six dewatering systems evaluated were: 1. conventional rotary vacuum filtration, 2. rotary precoat vacuum filtration, 3. pressure filtration, 4. porous bed filtration, 5. thermal spray drying, and 6. centrifugation.

No single dewatering system was found best for all acid mine drainage sludges. On the basis of cost, the most promising acid mine drainage sludge dewatering techniques appear to be centrifugation, rotary vacuum filtration and rotary precoat vacuum filtration.

17a. *Descriptors*

Acid Mine Drainage* neutralization* sludge* freezing, flocculation, centrifugation

17b. *Identifiers*

filter aid* vacuum filtration* pressure filtration*

17c. *COWRR Field & Group* Ø5D

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