EPA-600/2-78-014
February 1978

Environmental Protection Technology Series

PILOT INVESTIGATIONS OF SECONDARY SLUDGE DEWATERING ALTERNATIVES

Industrial Environmental Research Laboratory
Office of Research and Development
U.S. Environmental Protection Agency
Cincinnati, Ohio 45268

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into nine series. These nine broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The nine series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies
6. Scientific and Technical Assessment Reports (STAR)
7. Interagency Energy-Environment Research and Development
8. "Special" Reports
9. Miscellaneous Reports

This report has been assigned to the ENVIRONMENTAL PROTECTION TECHNOLOGY series. This series describes research performed to develop and demonstrate instrumentation, equipment, and methodology to repair or prevent environmental degradation from point and non-point sources of pollution. This work provides the new or improved technology required for the control and treatment of pollution sources to meet environmental quality standards.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

EPA-600/2-78-014
February 1978

PILOT INVESTIGATION OF SECONDARY SLUDGE DEWATERING ALTERNATIVES

by

Reid A. Miner
Duane W. Marshall
National Council of the Paper Industry
for Air and Stream Improvement, Inc.
Central-Lake States Regional Center
Kalamazoo, Michigan 49008

Isaiah Gellman
National Council of the Paper Industry
for Air and Stream Improvement, Inc.
New York, New York 10016

Grant No. R-804019-01-0

Project Officer

Victor J. Dallons
Industrial Pollution Control Division
Food and Wood Products Branch
Corvallis, Oregon 97330

INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

DISCLAIMER

This report has been reviewed by the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report concerns evaluating and comparing various sludge dewatering devices for use on biological sludges produced in the treatment of pulp and paper mill wastes. The information and comparisons developed should be useful to designers of pulp and paper waste and treatment systems to evaluate which sludge dewatering technology and which ultimate disposal scheme will be most applicable to the site in question. For further information about this report contact the Food and Wood Products Branch of the Industrial Pollution Control Division.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati

ABSTRACT

A pilot investigation of biological sludge thickening and dewatering alternatives, including pressure filtration, precoat vacuum filtration, filter belt pressing, capillary suction dewatering, gravity filtration, centrifugation, and ultrafiltration has been conducted on waste activated sludge resulting from the treatment of waste water from an integrated bleached kraft-fine paper mill. Based upon a criterion of attainable cake consistency, three levels of performance are indicated: (1) pressure filtration and precoat vacuum filtration generating the driest cakes, (2) filter belt pressing yielding intermediate cake consistencies, and (3) gravity filtration, centrifugation, and ultrafiltration resulting in relatively low cake consistencies. In general, performance has been found to be severely affected by changes in feed sludge consistency, the amount of sludge conditioning, and the sludge's specific resistance to filtration. The type and amount of sludge conditioning required has been shown to be extremely variable, depending upon the dewatering technique being employed, the level of performance being expected of it, and the consistency and nature of the sludge being dewatered.

CONTENTS

FIGURES

FIGURES
(continued)

TABLES

TABLES
(continued)

ACKNOWLEDGMENTS

The authors gratefully acknowledge the contributions of the following individuals and organizations in the conduct of this study.

P.H. Glatfelter Company	Mr. Phillip Hershey Dr. Neal Carter Mr. Larry Metzger Dr. Robert Olsen Mr. Dean Hoover Mr. Bill Higgs Mr. Bob Foti
Packaging Corporation of America	Mr. Roger Smithe
Johns-Manville Company	Mr. G.R. Bell Mr. Alan Wirsig
P-K Associates	Mr. Peter Kaye Mr. Doug McCord
Permutit Company	Mr. James Walzer Mr. Bill Coon
Sharples Division of Pennwalt Corporation	Mr. Dick Moll Mr. John McKenna Mr. Charles Hetrick Mr. Jack Malcomson
Tait-Andritz Corporation	Dr. John Leppitsch Dr. Joseph Bauer

In addition, the efforts of National Council for Air and Stream Improvement technical assistants Rick Hartman, Jane McClary, Dana Marks, Jeff Marks, Dirk Swinehart, Paul Turpin, and Tom Wistar are gratefully acknowledged.

The contributions of Mr. Russell O. Blosser, Assistant Technical Director of the National Council for Air and Stream

Improvement, and the secretarial assistance of Ms. E. Kavelman and Ms. C. Clark are especially appreciated.

Finally, the support of this study by the Research and Development Office of the United States Environmental Protection Agency, and the assistance provided by the Project Officer, Mr. V. Dallons, is gratefully acknowledged.

SECTION 1

INTRODUCTION

The process of wastewater treatment is primarily one of concentrating or converting potential pollutants into dilute slurries of solids, commonly called sludge, which require disposal. Because the costs of disposing of these solids are often strongly dependent upon the amount of water associated with them, most sludge handling systems include a dewatering stage. Due to the nature of the solids involved, sludges vary widely in their responsiveness to dewatering, with biological sludge being among the most difficult to dewater.

The most common method for dewatering biological solids in the pulp and paper industry has been in combination with primary solids on vacuum filters or centrifuges (1). This technique has been generally successful due to (a) the fibrous nature of much of the industry's primary sludge and (b) the existance of favorable ratios of primary-to-secondary solids, normally 5:1 to 3:1. However, there are a growing number of mills finding it advantageous or necessary to dewater sludges composed primarily or entirely of biological solids.

For instance, the sulfite and NSSC segments of the industry are characterized by high BOD generation rates relative to suspended solids losses resulting in primary-to-secondary sludge ratios of from 1:1 to 1:3. Recycled paperboard mills are able, in many instances, to reuse primary solids in the production process making them unavailable for admixing with secondary sludge. Even in those segments of the industry where the primary sludges have been typically fibrous in nature and the ratios of primary-to-secondary solids have been favorable, intensified efforts at minimizing fiber losses have resulted in diminished quantities of increasingly difficult-to-dewater primary sludge being available for admixing. Other mills are considering separate dewatering of primary and secondary sludges to preserve the superior dewaterability of their primary sludge, or protect by-product opportunities for either of the sludges. The sludges resulting from these sludge handling and dewatering practices have generally not been amenable to conventional vacuum filtration, decanter centrifugation, or V-pressing. Concurrent with the generation of more difficultly dewatered sludges, many mills are beginning to utilize final disposal methods that favor the

the generation of cakes with consistencies beyond the capabilities of applicable conventional dewatering technologies.

In general then, the industry's final disposal requirements, and changing sludge characteristics clearly indicate the need for dewatering technologies with capabilities beyond those normally associated with vacuum filtration, decanter centrifugation, and V-pressing.

This report describes the findings of a field investigation in which eight pilot dewatering devices were operated on biological sludge from the activated sludge process.

SECTION 2

CONCLUSIONS

All of the units investigated were capable of dewatering this mill's waste activated sludge.

The pressure filter and precoat vacuum filter generated the driest cakes (25 to 40 percent solids) making them likely alternatives where incineration, or dry cake for landfilling or hauling are involved in final disposal. The possible effects of handling and disposing of the nonsludge fraction of these cakes require consideration.

Pressure filter performance on the sludge studied was not enhanced by higher operating pressures (13 atmospheres v. 7 atmospheres). Additional industry experience suggests lower operating pressures to be adequate for other types of sludges as well.

The filter belt presses generated cake consistencies approaching 20 percent solids. Relatively low power and maintenance costs are likely to be associated with these units. The filter belt presses generated the driest cakes of the units not utilizing inorganic conditioning, suggesting them as a likely alternative where inorganic contamination of the sludge is not desired.

The gravity filter and centrifuge generated cakes at 8 to 10 percent maximum consistency, indicating probable applications in prethickening for pressure filtration, digestion or heat treatment, or where land disposal of semi-fluid sludge is envisioned. The gravity filter offers very low power costs while the centrifuge offers no sludge conditioning costs.

The ultrafilter was capable of thickening sludge to 7 percent consistency. However, the high power costs associated with overcoming the pressure drop through the system suggest that a different membrane configuration would be required to make the process feasible.

There were several units that consistantly operated at solids recoveries of 99 percent or better suggesting their applicability where the liquid fraction solids levels had to be low. These units were the pressure filter, the precoat vacuum filter, the gravity filter and the ultrafilter.

Artificially increasing the specific resistance of the unconditioned sludge from a range of 50 to 200 x 10^7 sec^2/gm to a range of 150 to 400 x 10^7 sec^2/gm had a substantial detrimental impact on the performance of those units which could handle the degraded sludge. Caution is, therefore, indicated in applying the data in this study directly to other sludges, particularly where knowledge of the specific resistance of the sludge in question is lacking.

Most of the units indicated a sensitivity to feed concentration and achieved or indicated significantly superior performance at sludge feed consistencies of 2 percent solids as compared to 1 percent solids.

Additional pulp and paper industry experience has shown both pressure filtration and filter belt pressing to be applicable to primary, secondary, and combined sludge dewatering. Performance levels are determined by the equipment configuration, the nature of the solids, the type and amount of sludge conditioning, and the feed consistency.

SECTION 3

RECOMMENDATIONS

This study has given rise to a number of issues related to, but not within the scope of, this investigation.

1. Documentation of the capabilities of these technologies on other pulp and paper industry sludges should be continued as more pilot and full scale data is generated throughout the industry.

2. As full scale pulp and paper industry experience on these technologies becomes increasingly available, an ongoing effort is warranted to (a) determine the reliability of manufacturer's performance estimates and (b) identify any chronic process limitations or maintenance problems that arise.

3. The information available on the capabilities of sludge dewatering technology requires delineation of an analysis framework conducive to the development of some optimal strategy for achieving specified ultimate disposal objectives at individual locations. Implicit in doing so is the availability of cost information adequate for comparisons.

4. Because the nature of the filtration media has been found in several instances to be a crucial variable in assuring acceptable performance, the need exists for a rational method of media selection based upon an understanding of the properties of the sludge-media interface.

SECTION 4

GENERAL DESCRIPTION

Recognizing the need for additional dewatering alternatives, especially for biological sludges, and realizing the importance of unbiased performance data in the selection of dewatering equipment, the National Council, with the cooperation of the Environmental Protection Agency, Cincinnati, Ohio, and P. H. Glatfelter Company in Spring Grove, Pennsylvania, and Packaging Corporation of America in Filer City, Michigan, conducted a study of the eight different emerging or novel dewatering alternatives listed in Table 1. All units except the ultrafilter were tested at P. H. Glatfelter Company, a bleached kraft mill producing 600 tons of fine papers per day. The sludge utilized was secondary clarifier underflow generated in the mill's contact stabilization activated sludge system. The sludge was characterized by consistencies of 0.5 to 1.5 percent solids, ash contents of 35 to 45 percent, a mixed liquor SVI (sludge volume index) of 100 to 200, and a specific resistance to filtration (2) (3) ranging from 50 to 200 x 10^7 sec^2/gm. The variability in the secondary sludge's resistance to dewatering is demonstrated in Figure 1. Because the specific resistance of the sludge was relatively low for biological sludge in general, efforts were made, when possible, to also test the equipment with sludge that had been physically degraded by either aging or pumping through a centrifugal pump. Sludges thus treated typically had a specific resistance of 200 to 1000 x 10^7 sec^2/gm suggesting that centrifugal pumping and sludge aging might contribute to decreased sludge dewaterability at other mills as well. Because of the variability in sludge characteristics from mill to mill, the data generated in this study are not necessarily directly applicable to sludges at other locations, particularly where a knowledge of the specific resistance of the sludge in question is lacking. Furthermore, the documented variability at an individual location suggests the importance of collecting sludge dewaterability data over a period longer than, but inclusive of, any pilot study undertaken.

TABLE 1. BIOLOGICAL SLUDGE DEWATERING ALTERNATIVES STUDIED

Technology	Manufacturer
Pressure filter	Netzsch
Precoat vacuum filter	Dorr-Oliver, pilot unit obtained through Johns-Manville Corp.
Filter belt press	Permutit
Filter belt press	Tait-Andritz
Capillary filter belt press	Prototype built by Westinghouse under EPA contract
Biological sludge deanter centrifuge	Sharples
Gravity filter	Permutit
Ultrafilter	Westinghouse modules utilized, assembled by the National Council

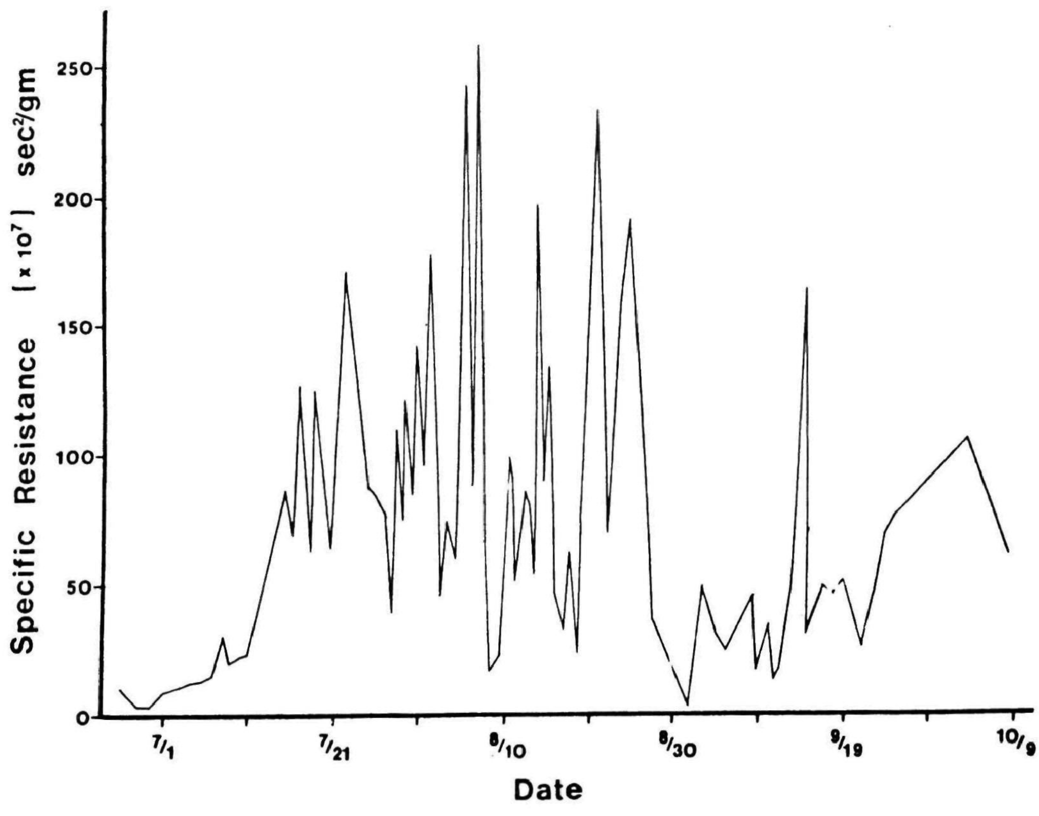


Figure 1. Variability of biological sludge dewaterability.

SECTION 5

PRESSURE FILTRATION INVESTIGATION

PROCESS DESCRIPTION

Pressure filtration is a batch dewatering process using a series of filter media-lined chambers formed by adjoining recessed plates. Sludge is forced into the chambers under pressure, the liquid passing through the filter media into collection troughs in the plates and away from the chambers. The solids meanwhile accumulate in the chambers until additional solids are entering at a very low rate. At this point the plates are separated and the cakes of accumulated solids are discharged. These filters can be operated in a precoat mode utilizing diatomaceous earth or flyash (or other suitable materials) to provide a layer of solids between the sludge solids and the filter media, to provide protection for the filter media and promote clean cake discharge. The operating variables for the pressure filtration process are listed in Table 2.

DESCRIPTION OF PILOT UNIT

The pressure filter used in this study was a three-chamber Netzsch filter with a capacity of 0.87 gallons (3.3 liters). The cakes generated on this unit were 10 in. (25.4 cm) square and 1 in. (2.54 cm) thick. The three chambers offered a total of 2.94 ft^2 (0.27 m^2) of filter area. The operating pressure was variable from 0 to 200 atmospheres (0 to 200 bar) and was maintained with a progressing cavity sludge pump in conjunction with an air compressor and pressurized feed tank as shown in Figure 2. The three different filter media that were used are described in Table 3.

OPERATION OF THE PILOT NETZSCH PRESSURE FILTER

In those cases when a diatomaceous earth precoat was required, it was applied at 0.13 lb of diatomaceous earth per ft^2 filter area (0.63 kg/m^2). This amount was determined through a procedure involving visual examination of the deposited precoat. A diatomaceous earth slurry at about 1 percent concentration was introduced to the unit between the pressure

TABLE 2. PRESSURE FILTRATION PROCESS VARIABLES

Independent variables
1. Chamber dimensions and characteristics
2. Media characteristics
3. Precoat utilization
4. Precoat characteristics
5. Operating pressure
6. Cycle time
7. Sludge consistency
8. Sludge conditioning
9. Nature of the sludge solids
Dependent variables
1. Cake consistency
2. Loading rate
3. Solids recovery

TABLE 3. CHARACTERISTICS OF PRESSURE FILTER MEDIA

Media A	- monofilament
	4:1 satin weave
	56 threads/inch x 56 threads/inch
	interthread opening - 0.37 mm x 0.29 mm
Media B	- monofilament
	4:1 satin weave
	46 threads/inch x 46 threads/inch
	interthread opening - threads tightly packed
Media C	- multifilament
	3:2 satin weave
	150 threads/inch x 35 threads/inch
	interthread opening - threads tightly packed

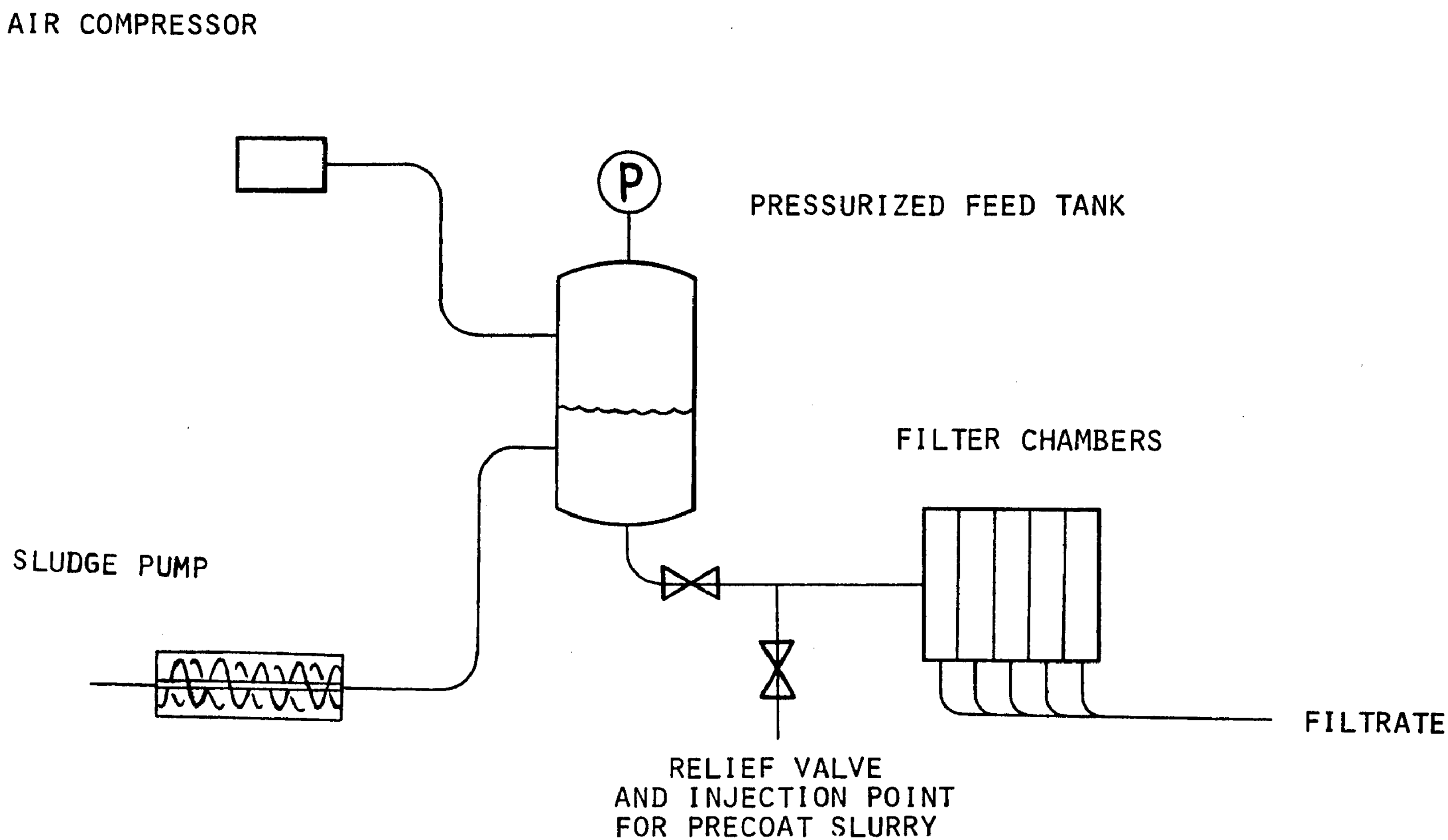


Figure 2. Netzsch pressure filter.

tank and the filter chambers at 0.5 to 1 gpm/ft^2 of area (20-40 l/min/m^2). The resulting filtrate was recirculated to the diatomaceous earth slurry feed vessel until the vessel contents were clear. This flow rate was found necessary to achieve even distribution of the diatomaceous earth. A somewhat lower rate would be suitable for less easily filterable precoat materials. After the diatomaceous earth slurry feed tank contents were clear, it was found necessary to pump air through the chambers until the majority of the water had been removed from the deposited precoat. This prevented the precoat from sliding to the bottom of the chambers prior to sludge's being introduced.

Test runs were preceded by adjusting the pressure regulators to the desired operating pressure. The test was initiated by turning on the feed pump and air compressor causing sludge to be forced into the filtration chambers. Stopwatch timing of the run was started when the first drop of filtrate was collected. The remainder of the test consisted of recording the volume of filtrate collected and the pressure attained as a function of time. A typical time/volume/pressure relationship is shown in Figure 3. Although in some circumstances the high initial rate of increase in pressure would indicate media blinding, in the study the fact that cake consistencies and cycle times were reasonable suggests that the high initial rate of increase in pressure is a result of an oversized feed pump in relation to the total chamber volume. Work by Martin and Hayden (4) suggests that the performance levels achieved with a rapid pressure increase are comparable to those attained under conditions of gradually increasing pressure.

After the prescribed length of cycle time, the pressure was relieved, the plates were separated and the cakes removed. The middle cake was weighed and three samples were taken from it: one from the upper and one from the lower right corners and one from halfway between the center of the cake and the middle of the right edge. The effective bulk cake consistency was defined as the average of these three to minimize the impact of consistency variation throughout the cake. Samples were also taken of the conditioned and unconditioned feed and the filtrate. Specific resistance determinations on the conditioned and unconditioned sludges allowed judgments as to the nature of the unconditioned solids and the effectiveness of the conditioning utilized. To aid in the assessment of the potential advantages of precoat utilization, the ease of cake discharge and media cleaning were noted.

RESULTS OF PRESSURE FILTRATION TESTS

Loading Rates

The rate at which the filter could handle solids depended upon the required cake consistency, the feed concentration, the

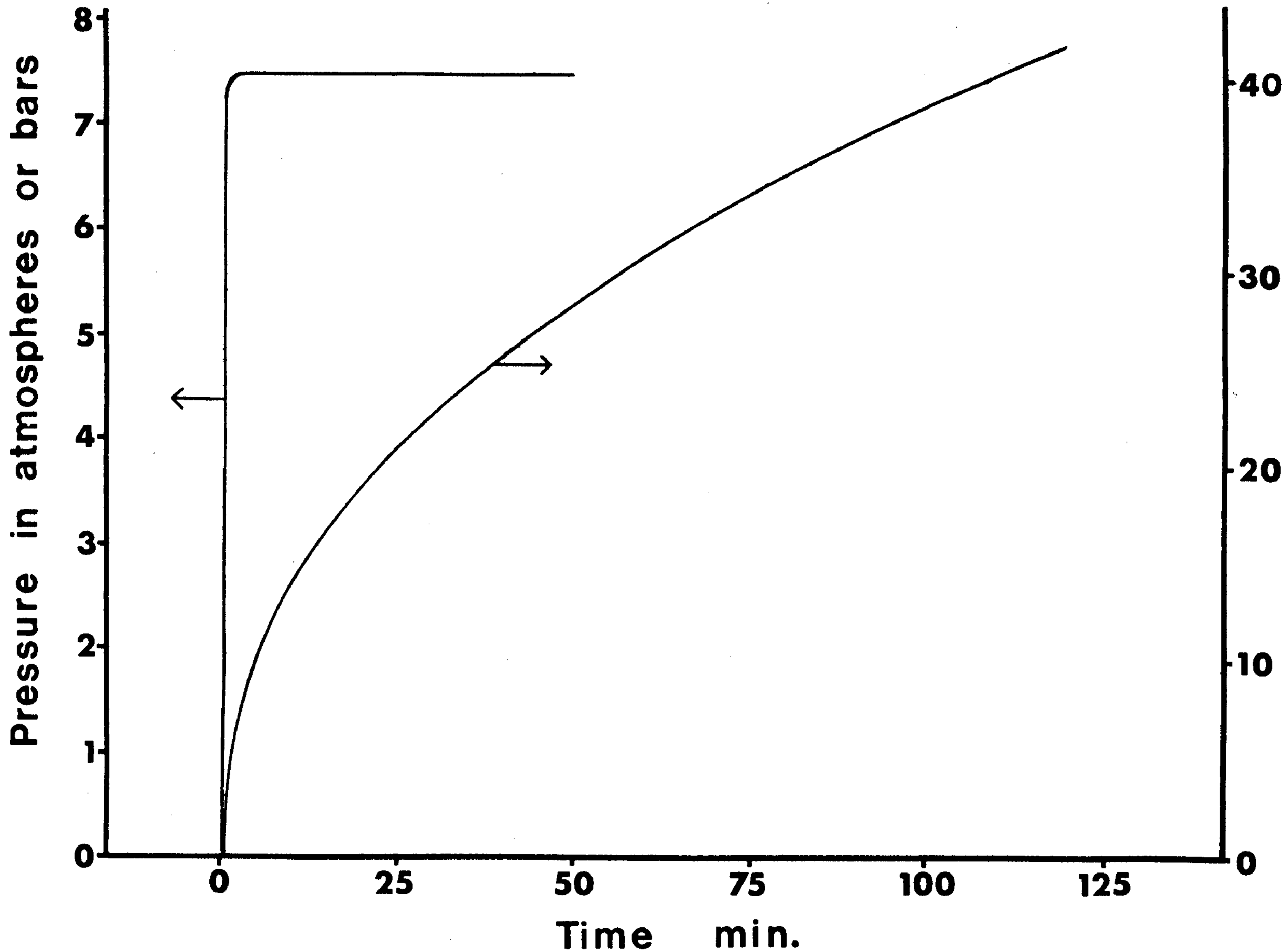


Figure 3. Typical pressure filtration cycle.

nature of the solids, and the type and amount of conditioning. Loading rates are related to cake consistencies through the cycle time and solids accumulation. Figure 4 shows that the bulk cake consistency increases with the total weight of solids deposited in the filter. The desired cake consistency is then achieved when sufficient cake solids are accumulated. The quantity of solids deposited at any particular time is calculated from the total flow and feed concentration. The total solids deposited per filter area divided by the cycle time is the pilot loading rate. However, because of the fact that areas around the feed ports are of lower cake consistency than the rest of the filter surface and because these areas represent a much larger proportion of the filtration surface in pilot than in full scale units, using the calculated pilot loading rates will result in values much lower than commonly encountered in full scale. On this particular pilot unit, the problem is further aggravated by a media-anchoring plate which obstructs a significant portion of the filtration area in the first of the three chambers. Because of the difficulties encountered in determining meaningful pilot loading rates, the values in Table 4 are the loading rates calculated at the effective cake consistency (defined in the previous section). These effective loading rates are determined by calculating the weight of solids that would be present in the filled press if the cakes were of uniform consistency using the following equation

$$\text{Loading Rate} = \frac{\text{Press Volume x Cake Density}}{\text{Press Area x Cycle Time}}$$

where the cake density is based upon the effective cake consistency as presented in Reference 5. As shown in Table 4, other things being equal, lower loading rates are required to obtain higher cake consistencies. The table also illustrates that the benefits associated with prethickening of pressure filter feed are substantial, allowing reductions in sludge conditioning costs, shorter cycle times, or higher cake consistencies. Physically degraded solids required significantly lower loading rates or more conditioning to achieve cake consistencies comparable to fresh waste activated sludge filter cakes.

Seven separate sets of runs were conducted at different times in the pilot program to document the magnitude of the advantages associated with higher operating pressures. The results, summarized in Table 5, provide little basis for using 13.5 atmospheres of pressure in the filtration of this sludge. Acceptable capacities were maintained in the operating pressure range of 6 to 8 atmospheres. One issue not addressed in this study was the effect of chamber depth upon loading rate and optimum operating pressure.

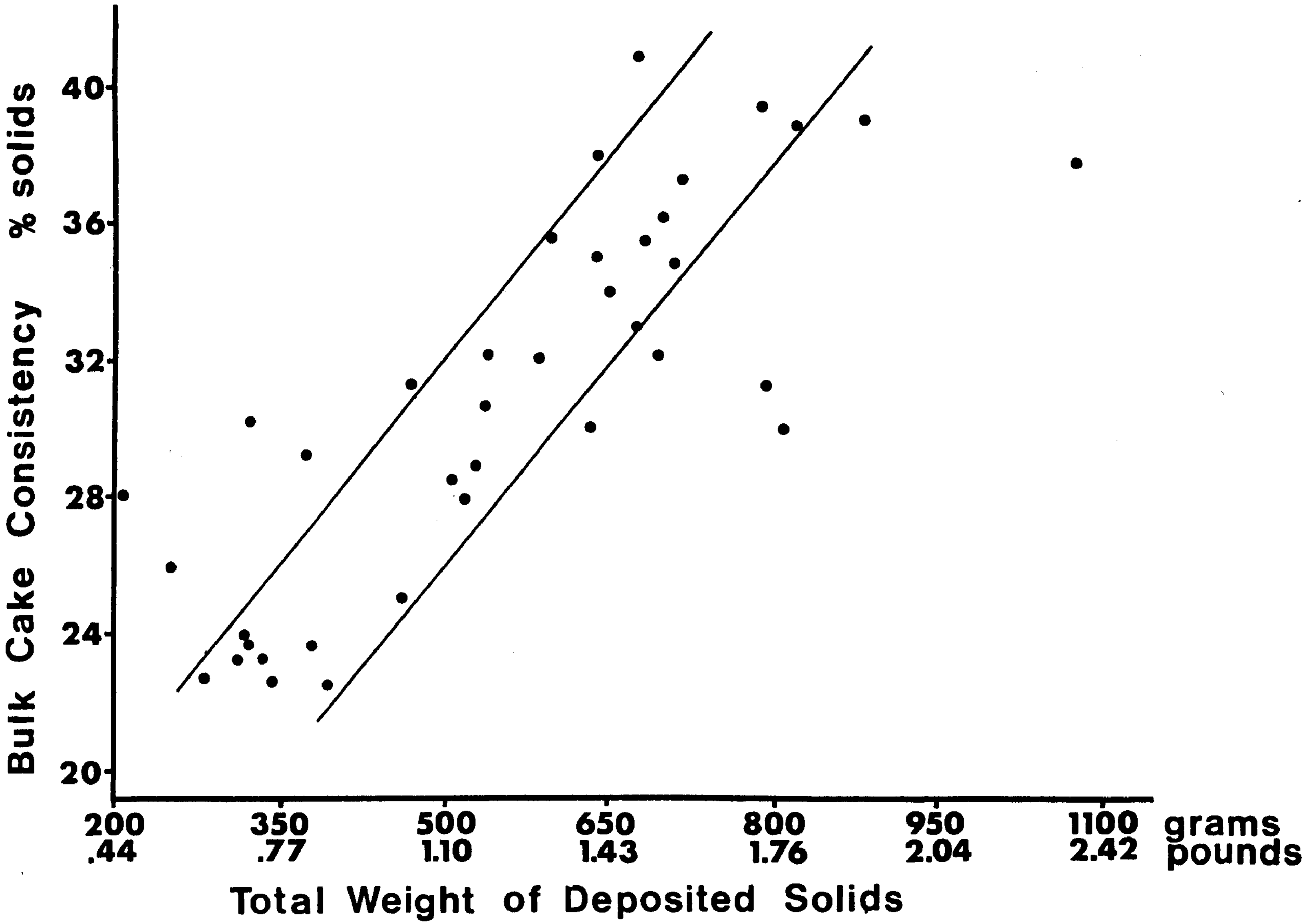


Figure 4. Relationship between bulk cake consistency and accumulated solids mass.

TABLE 4. PRESSURE FILTER PERFORMANCE SUMMARY

Feed consistency % solids	Bulk cake consistency % solids	Cycle time requirement hr	Conditioning requirement	Loading rate* $\#/ft^2/hr$ 1" cake	Loading rate* #/gal/hr
FRESH WASTE ACTIVATED SLUDGE					
0.7-1.00	26-32	1.00-1.50	25-30% lime & 7-8% $FeCl_3$	0.81-0.54	2.72-1.82
0.7-1.00	32-38	2.00-3.50	35-40% lime & 8-10% $FeCl_3$	0.49-0.28	1.64-0.94
1.1-1.35	26-32	0.75-1.00	25-30% lime & 7-8% $FeCl_3$	1.08-0.81	3.63-2.72
1.1-1.35	32-38	1.00-2.00	25-30% lime & 7-8% $FeCl_3$	0.97-0.49	3.28-1.64
1.6-1.70	26-32	0.50	25% lime & 7% $FeCl_3$	1.62	5.44
1.6-1.70	32-38	0.50-0.75	25% lime & 7% $FeCl_3$	1.94-1.29	6.56-4.37
1.0-2.00	38-40	3.00-3.50	25-30% lime & 7% $FeCl_3$	0.37-0.32	1.25-1.07
AGED OR HIGHLY SHEARED WASTE ACTIVATED SLUDGE					
0.6-0.90	26-32	2.00-3.50	35-45% lime & 8-10% $FeCl_3$	0.41-0.23	1.39-0.78
1.0-1.30	26-32	0.75-1.00	35-40% lime & 7-8% $FeCl_3$	1.08-0.81	3.65-2.74
1.0-1.30	32-38	1.00-2.00	35-40% lime & 7-8% $FeCl_3$	0.97-0.49	3.28-1.66
1.4-1.60	26-32	0.75-1.00	25-30% lime & 6-7% $FeCl_3$	1.08-0.81	3.65-2.74
1.4-1.60	32-38	1.00-2.00	25-30% lime & 6-7% $FeCl_3$	0.97-0.49	3.28-1.66
1.0-2.00	38-40	3.00-3.50	35-40% lime & 7% $FeCl_3$	0.37-0.32	1.25-1.08
2.1-2.70	26-32	0.50-0.75	15-20% lime & 4-6% $FeCl_3$	1.62-1.08	5.47-3.65
2.1-2.70	32-38	1.00-1.50	15-20% lime & 4-6% $FeCl_3$	0.97-0.65	3.28-2.20
11.2**	26-32	0.25-0.50	30% lime & 7% $FeCl_3$	3.24-1.62	10.95-5.47
11.2**	32-38	0.50-0.75	30% lime & 7% $FeCl_3$	1.94-1.29	6.56-4.36

*Based upon cake densities of 68, 70 and 72 pounds per wet cubic foot for cakes of 30, 35, and 39 percent solids respectively (Reference 5)

**Only one test run attempted with centrifugally thickened, sheared, waste activated sludge

TABLE 5. EFFECTS OF OPERATING PRESSURE UPON PRESSURE FILTRATION

Data set	Date	Pressure atm. or bar	Cycle time hr	% Lime /% $FeCl_3$ conditioning	Bulk Cake consistency % solids	Weight of accumulated solids gm	Time to collect 200 grams of solids min
1	9/4	5.1	2.0	16/4	22.7	277	60.0
1	9/4	7.5	2.0	11/3	24.4	320	32.0
1	9/4	13.5	2.0	15/4	20.5	257	70.0
2	9/5	5.1	2.0	29/7	33.0	675	7.0
2	9/5	7.5	2.0	25/6	35.5	685	6.0
2	9/5	13.5	2.0	26/6	35.7	595	8.0
3	9/8	7.5	1.0	35/8	32.2	587	3.1
3	9/8	13.5	1.0	34/8	30.6	540	5.5
4	9/22	6.8	2.0	24/6	23.3	318	38.0
4	9/22	10.6	2.0	24/6	23.3	339	44.0
4	9/22	13.5	2.0	29/7	24.4	309	22.0
5	9/23	6.8	2.0	20/5	23.9	318	34.0
5	9/23	10.5	2.0	20/5	23.6	378	24.0
5	9/23	13.5	2.0	20/5	22.5	347	28.0
6	9/24	6.8	1.2	38/10	34.7	710	7.0
6	9/24	13.5	3.3	32/8	39.3	797	8.0
7	10/1	6.8	1.6	32/8	30.4	630	6.6
7	10/1	13.5	3.3	31/8	32.2	541	16.3

Cake Consistencies

The higher cake consistencies shown in Table 4 were associated with (a) fresh waste activated solids, (b) large dosages of sludge conditioning, (c) thick feed, and (d) long cycle times. When adjusted for admix content (by using the difference between the consistencies of unconditioned and subsequently conditioned feed) the effective bulk cake consistencies in Table 4 are reduced to the corrected values shown in Table 6.

TABLE 6. CORRECTION OF CAKE CONSISTENCIES FOR ADMIX CONTENT

Bulk Cake consistency % solids	% Lime	Corrected cake consistency % solids
26-32	15-20	22-31
26-32	25-30	20-29
26-32	36-40	17-26
32-38	15-20	27-38
32-38	25-30	25-34
32-38	35-40	22-31
38-40	25-30	30-37
38-40	35-40	27-32

Solids Recovery

If two conditions were satisfied, the solids recoveries attained on the pressure filter remained in excess of 97 percent, usually being equal to or greater than 99 percent. First, the correct media was required. Media C performed best in this regard. Second, the sludge had to be properly conditioned. When these conditions were satisfied, solids recoveries were always in excess of 97 percent, even when filtration rates were unacceptably low. This suggests that the conditioning requirements for acceptable recoveries are less than those for adequate loading rates. Thus, solids recovery is assured when acceptable throughputs are maintained.

Conditioning

One aspect of the pressure filter pilot study was the evaluation of alternative conditioning techniques, the results of which are shown in Table 7. Based upon this conditioning evaluation, which showed lime and ferric chloride to be the type of conditioning which gave the most consistant results, it was decided to conduct the study of other operating variables using

TABLE 7. ALTERNATIVE CONDITIONING TECHNIQUES EVALUATED FOR PRESSURE FILTRATION

Conditioning	Results
1. Lime + ferric chloride	Consistantly acceptable results*
2. Lime	Unsuccessful at 1/3:lime/sludge
3. Lime + Betz 1260	Unsuccessful at 2/3:lime/sludge
4. Lime + Percol 725	Unsuccessful at 2/3:lime/sludge
5. Flyash	Unsuccessful at 2/1:flyash/sludge
6. Flyash + lime + ferric chloride	No improvement over lime + ferric chloride
7. Flyash + Herculoc 812.3	Unsuccessful at 1/1:flyash/sludge
8. Flyash + Percol 140	Unsuccessful at 1/1:flyash/sludge
9. Betz 1260	Unsuccessful
10. Hercufloc 812.3	Unsuccessful
11. Percol 722	Unsuccessful
12. Percol 140	Unsuccessful
13. Diatomaceous earth admix + Hercufloc 812.3	Successful at 3/1:sludge/diatomaceous earth
14. Lime mud ($CaCO_3$)	Unsuccessful at 1/3:mud/sludge
15. Lime mud ($CaCO_3$) + Percol 140	Unsuccessful
16. Repulped broke (reslurried paper)	Successful at 1/1:sludge/broke
17. Bark fines (85% by weight between 20 and 100 mesh)	Unsuccessful at 1/3:bark/sludge

* successful runs were those resulting in the formation of removable cakes in cycle times comparable to those shown in Table 4.

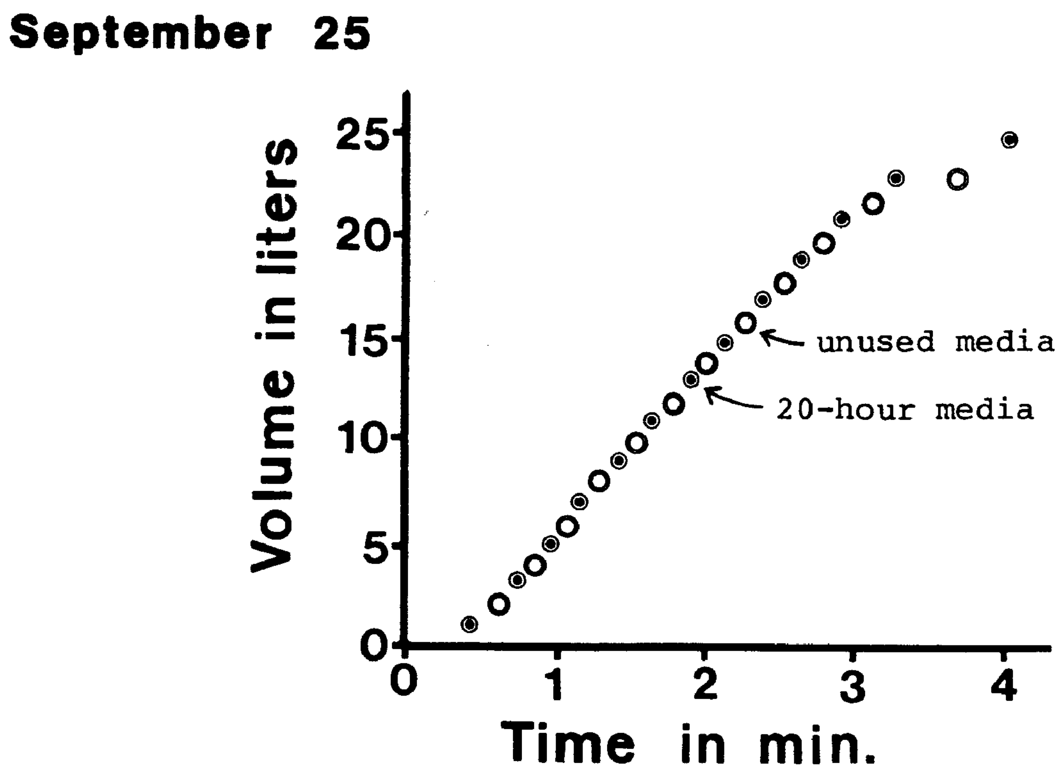
lime and ferric chloride for conditioning. As shown in Table 4, the amount of lime and ferric chloride required is higher for (a) degraded sludge solids, (b) higher cake consistencies, (c) shorter cycle times, and (d) less concentrated feed.

Utilization of Precoat

Another aspect of the pressure filtration study was the investigation of the advantages associated with precoat utilization. Based upon the cake discharge and media cleanup characteristics, it was shown that (a) precoat provided insurance that the media would be protected and (b) cake discharge would be acceptable even when the filtration characteristics of the sludge resulted in wet, structurally weak cakes. The advantages associated with the utilization of precoat when dry, tough cakes were produced were much more difficult to document than when wet cakes were generated.

To document the advantages of using precoat, two identical sets of filter cloth "B" were obtained for the press. One was utilized for a total of 29 operating hours without precoat, the other was not used for sludge dewatering. Periodically both sets of media were used to filter diatomaceous earth slurries. The easily filtered diatomaceous earth slurries allowed the identification of any significant blinding in the used media. For approximately 20 of the 29 operating hours, the used media were shower cleaned between runs, while no between run cleaning occurred during the last 9 hours of operation. Although the media used without precoat was visibly soiled with sludge solids, the diatomaceous earth filtration rates were not significantly different for the used and unused media, as shown in Figure 5. The length of time over which the precoat utilization study was conducted was probably insufficient to identify any long-term blinding tendencies that would be minimized through precoat utilization. However, the results of this study suggest that precoat may not be a universal requirement for the pressure filtration of waste treatment sludges.

Another aspect of the precoat utilization study was the evaluation of alternative precoat materials, specifically flyash and lime mud. Within the previously mentioned constraints upon determining long-term blinding problems, the flyash performed satisfactorily, distributing evenly, allowing high filtration rates, providing clean cake discharge and facilitating media cleanup. Lime mud ($CaCO_3$) also satisfactorily protected the media from sludge solids but it was not determined whether or not filtration rates were reduced from diatomaceous earth levels (as was to be the case in precoat vacuum filtration).



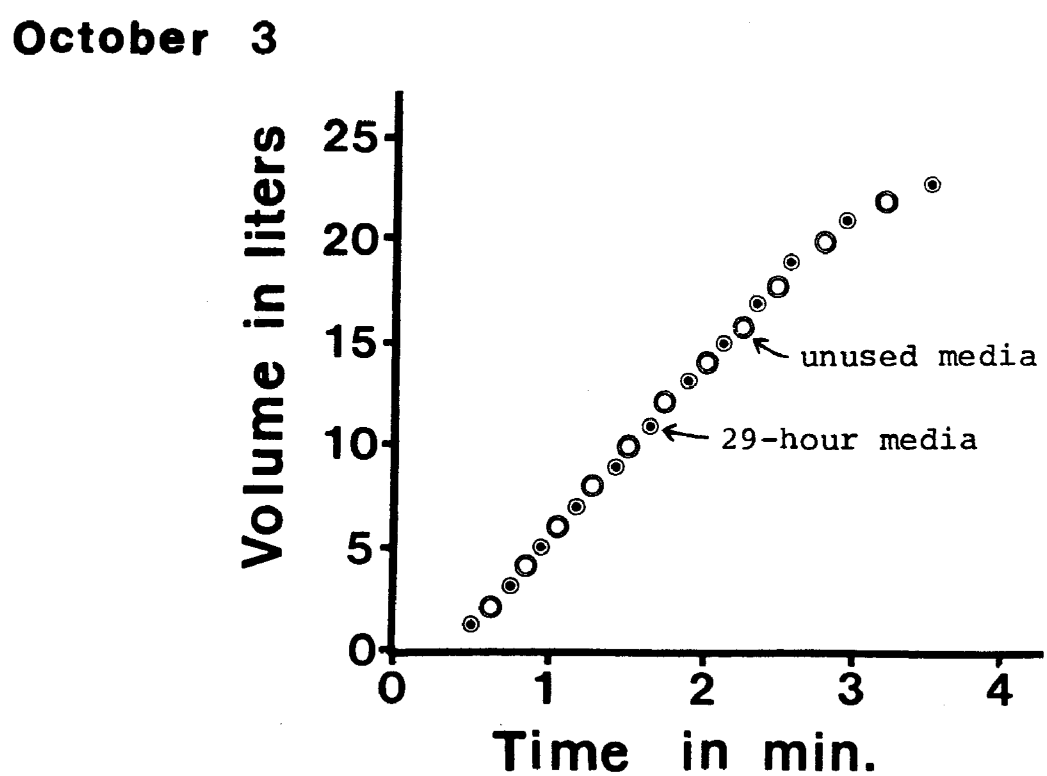


Figure 5. Diatomaceous earth slurry filtration rates through unused and used, nonprecoated media.

Overall Performance of the Pressure Filter

The pressure filter was capable of operating at a variety of performance levels. Table 4 demonstrates the interrelationship between the various process and sludge variables. The optimum set of operating conditions could vary greatly depending upon final disposal requirements and other individual circumstances. Cake consistencies of from 26 to 40 percent solids were attainable at conditioning requirements of 15 to 45 percent lime supplemented with 4 to 10 percent ferric chloride over cycle times of 15 minutes to 3.5 hours. The most important operating variables were found to be the feed concentration, the degree of conditioning and the nature of the solids. Operating pressures of from 5.1 to 13.5 atmospheres gave comparable results. The issue of the need for precoat remained unresolved, the data suggesting that the use of precoat may not be a universal requirement.

SECTION 6

PRECOAT VACUUM FILTRATION INVESTIGATION

PROCESS DESCRIPTION

Vacuum filtration is presently a commonly employed dewatering technique used for both primary and combined sludges. However, the application of rotary vacuum filtration to biological or other difficult-to-filter sludges has been limited chiefly by the inability to remove very thin sludge cakes from filter media in a manner which minimizes the potential for media blinding. The utilization of a filter precoat makes the removal of very thin cakes possible while renewing the filtering surface to a nonblinded condition. To accomplish this, the deposited sludge solids are removed from the filtration surface together with a thin layer of the precoat material. The advent of the rotating knife-doctoring assembly shown in Figure 6 has in many instances made it possible to reduce precoat consumption below the levels attainable with fixed knife-doctoring systems. The rotary precoat vacuum filtration process differs from conventional vacuum filtration in that (a) the filtration surface is a precoat material, and (b) the doctor blade assembly, be it fixed or rotating, advances toward the drum surface. The variables of importance to precoat vacuum filtration are shown in Table 8.

DESCRIPTION OF PILOT EQUIPMENT

The pilot rotary precoat vacuum filter provided by Johns Manville Corporation for use in this study, was a 36-in. (91 cm) diameter by 6-in. (15.2 cm) wide Dorr Oliver drum vacuum filter offering from 3.62 ft^2 to 5.85 ft^2 (0.34 to 0.54 m^2) of filtering surface, depending upon the width of the precoat face being doctored. The doctoring mechanism was of the rotating knife variety depicted in Figure 6.

OPERATION OF THE FILTER

Precoat was applied in a slurry at roughly 5 to 7 percent solids, at maximum vacuum (10 to 20 in. [250 to 500 mm] Hg, depending upon the precoat type), at 10 to 25 percent drum submergence, and at a drum speed of 1 to 2 RPM.

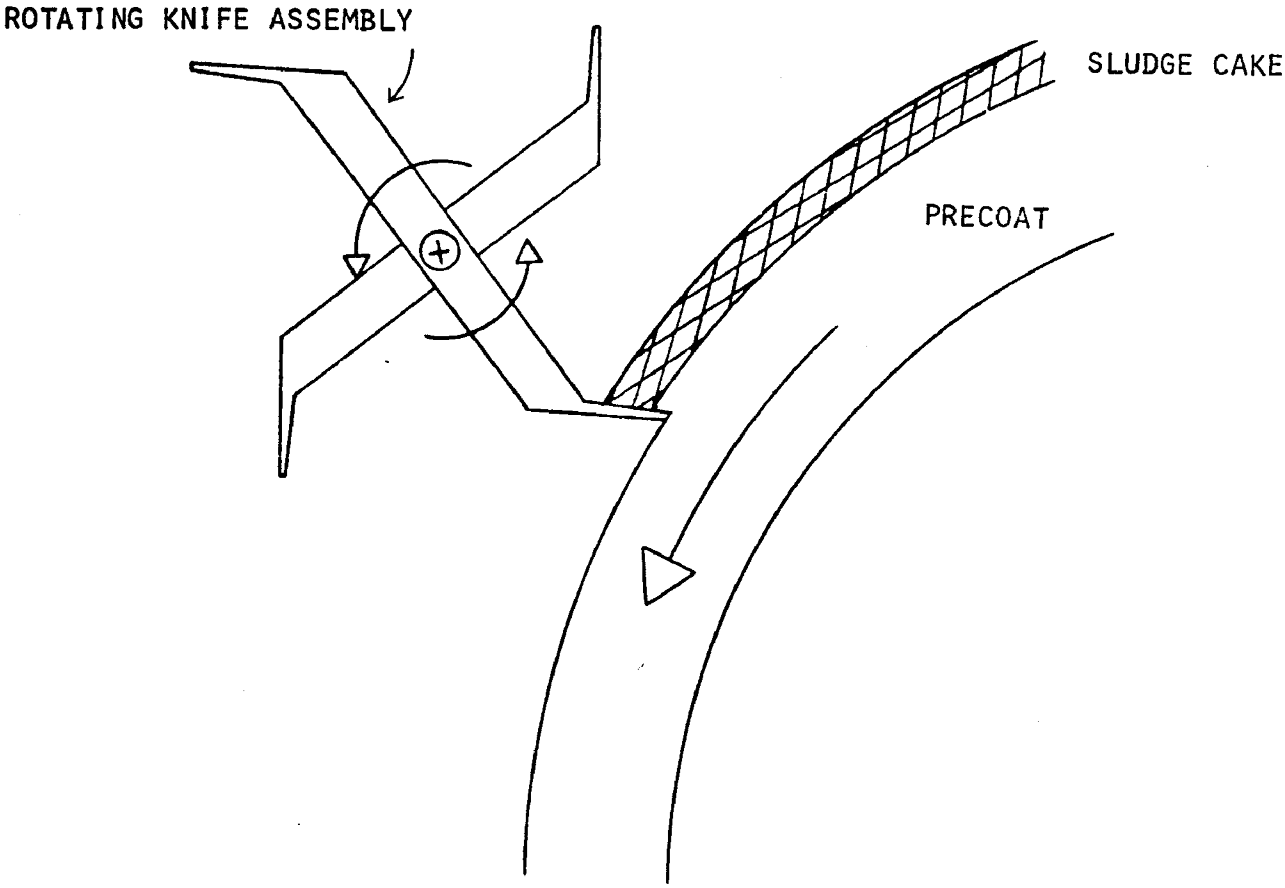


Figure 6. Rotating knife doctor blade assembly.
(SLUDGE CAKE THICKNESS AND CUT DEPTH EXAGGERATED)

TABLE 8. PRECOAT VACUUM FILTRATION PROCESS VARIABLES

Independent variables
1. Precoat characteristics
2. Type of doctoring assembly
3. Rotational speed of rotating knife doctoring assembly when utilized
4. Rate of knife advance (precoat consumption)
5. Drum speed
6. Drum submergence
7. Vacuum intensity
8. Feed consistency
9. Sludge conditioning
10. Nature of sludge solids
Dependent variables
1. Loading rate
2. Cake solids
3. Solids recovery

Once the precoat had been applied, sludge was introduced into the vat to seal the precoat surface, providing sufficient vacuum to assure precoat adherance to the filter face. After the precoat surface had been sealed, the precoat was allowed to stabilize for 30 minutes to 1 hour assuring that any precoat shrinkage occurred before the test runs were conducted. This stabilization was accomplished by operating the filter at maximum vacuum and a knife advance of 1 mil (25.4 microns) per drum revolution. After the precoat was stabilized, the vat level and drum speed were adjusted to desired values, and the process of optimizing precoat consumption was begun.

An initial knife advance of from 1 to 2 mils (25.4 to 50.8 microns) per revolution was selected based upon the results obtained during previous runs. Initially, 1 hour of complete cake removal and uninhibited filtration rates were used as the criteria for an adequate knife advance rate, but as the operating personnel became familiar with the unit, and as the appropriate advance rates became more evident, 30 minutes were felt to be sufficient indication. Knife advance rates were reduced until unacceptable cake removal was encountered indicating that the next highest rate was optimum. Once the optimum knife advance was found, samples were taken of the sludge in the filter vat, the cake being generated, and occasionally the filtrate. In addition, the filtrate rate was measured, generally over a period of 3 to 10 minutes, to minimize the possible influence of individual precoat laminations, which have been identified as offering greater resistance to flow than the bulk precoat material (6). In conjunction with each filtrate rate, the effective width of the filter face (the width being doctored) was measured. A specific resistance test was conducted on each vat sample to permit identification of changes in sludge dewaterability.

PRECOAT VACUUM FILTER RESULTS

Solids Loading Rates

The maintenance of acceptable solids loading rates was found to require (a) a rate of knife advance suitable for maintaining a filtering surface free of sludge solids and (b) the utilization of a precoat with suitable water permeability. The recommendation of the precoat suppliers, based upon their previous experience and precoat filter leaf tests conducted at the site, was Hyflo* diatomacous earth.

The solids loading rates attainable with the vacuum filter with a diatomaceous earth precoat were found to be influenced by (a) the specific resistance of the feed, (b) the concentration of the feed, (c) the drum speed, and (d) the degree of drum submergence. Figure 7 demonstrates the effect of feed consistency, drum speed, and submergence upon solids loading rates achieved on sludges with specific resistances of less than 200 x 10^7 sec^2/gm. The effect of feed consistency upon solids loading rate was quite pronounced, resulting in 80 to 300 percent increases in loading rates when feed consistencies were increased from 0.8 percent to 2.0 percent solids. For sludges with a specific resistance below 200 x 10^7 sec^2/gm, variations in specific resistance did not usually result in significant variations in loading rate. However, upon those occasions when the specific resistance of the sludge exceeded 200 x 10^7 sec^2/gm, severe reductions in loading rates were noted. Although the data are sparce for sludges with specific resistances greater than 200 x 10^7 sec^2/gm, it is nonetheless clear that the loading rates shown in Figure 8 fall significantly below their lower specific resistance counter parts.

Two separate attempts were made to determine the effect of vacuum intensity on solids loading rates. In one case, the highest solids loading rate achieved was at a vacuum of 10 in. (250 mm) of Mercury while in the second instance, the lowest loading rate was associated with the same vacuum. Thorough examination of the two sets of data has not provided an explanation for this apparent contradiction.

* Product of Johns-Manville Corporation

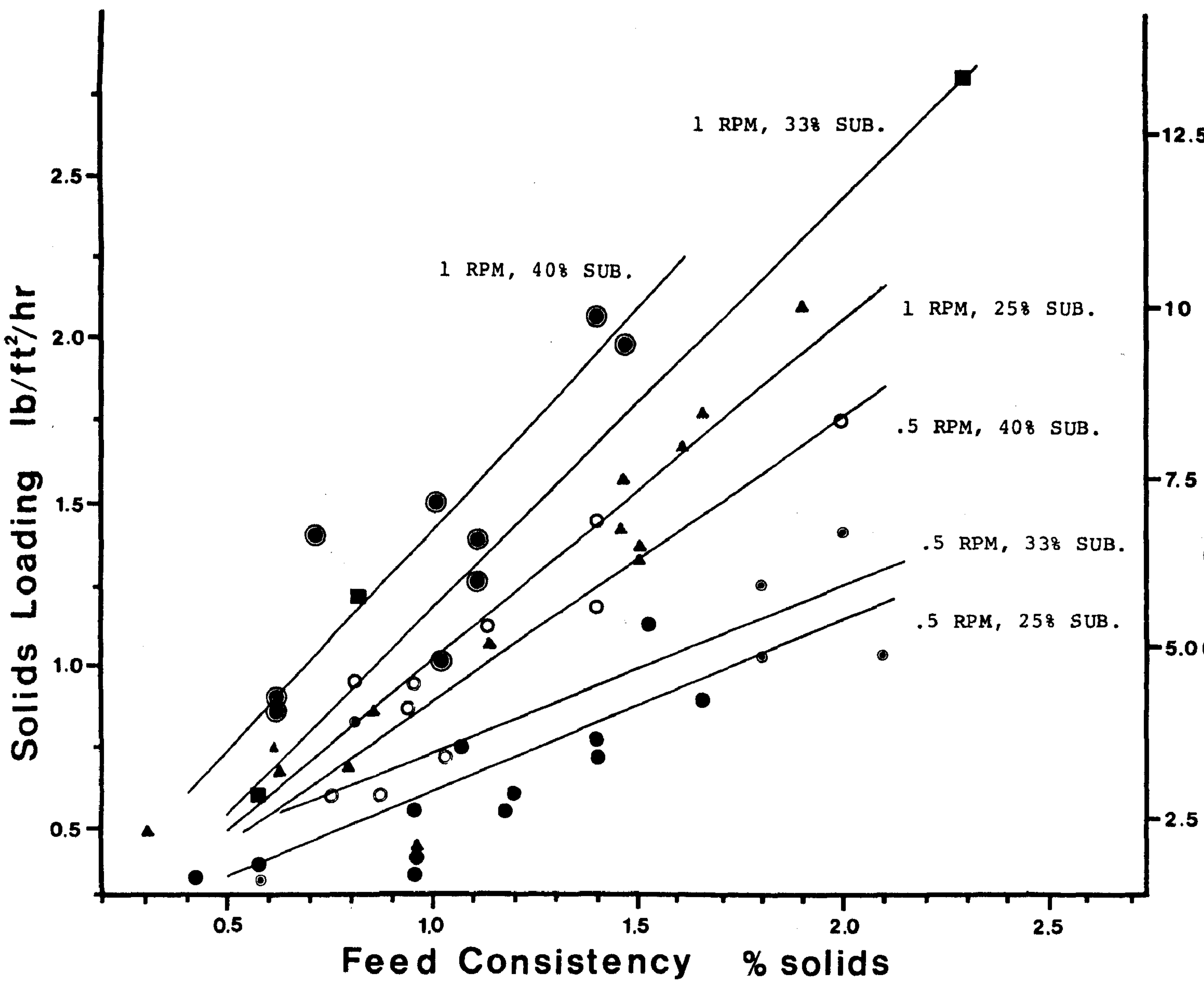


Figure 7. Precoat vacuum filter solids loading rates.

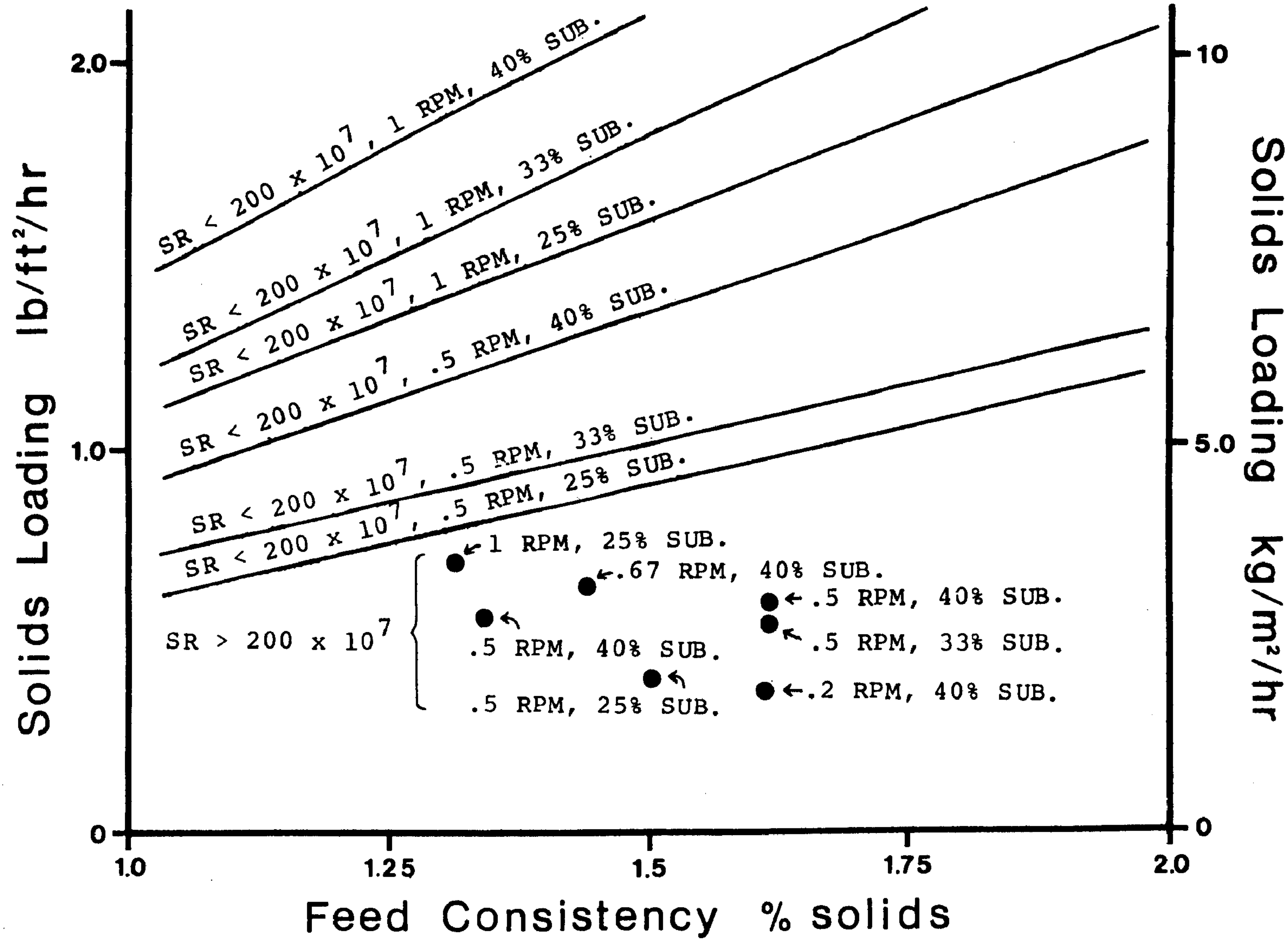


Figure 8. Loading rates attained on high specific resistance (SR) sludges.

Cake Consistencies

The cake consistencies attained with a diatomaceous earth precoat, after being corrected for precoat content, were found to be influenced primarily by drum speed, drum submergence and vacuum intensity. Figure 9 constructed from all data collected with diatomaceous earth precoats on sludges with a specific resistance of less than 200 x 10^7 sec^2/gm and with vacuum intensities of 20 in. (250 mm) of mercury or greater, demonstrates that slow drum speeds and low drum submergences contributed to higher cake consistencies. In addition, the ability of higher vacuums to produce drier cakes was quite pronounced as shown in Figure 10. The cake consistencies attained on sludges with a specific resistance greater than 200 x 10^7 sec^2/gm were generally about 5 percent solids content below those typical of sludges with a lower specific resistance.

Precoat Vacuum Filter Solids Recoveries

The solids recoveries attained with diatomaceous earth precoats (Hyflo* and Celite 545*) were measured on 11 different occasions. All 11 values fell between 98 and 100 percent solids recovery, with 8 of the 11 values being in excess of 99.9 percent. Because the knife advances associated with the use of Celite 545* were frequently in the 2 to 3 mil/revolution range (75 to 100 micron/revolution) the acceptable filtrate qualities and loading rates attained do not necessarily suggest Celite 545* as an acceptable precoat material in this application. Hyflo* is the grade of diatomaceous earth usually used in this application. The solids recoveries associated with the use of flyash precoats were more variable. Of the 7 determinations, 4 were in excess of 99.8, the lowest value of the 7 being 97.3 percent. During one day's testing with flyash precoat, the filtrate solids levels decreased from 800 to 70 parts per million throughout the day suggesting that the precoat was being penetrated and filled with sludge solids. However, this gradual plugging did not have a noticeable effect upon solids loading rates.

Precoat Consumption

The minimum precoat consumption that assured consistantly acceptable performance was determined to be 1 mil (25.4 microns) of Hyflo* precoat per drum revolution. At drum speeds greater than 1 revolution per minute, it was not possible to attain consistantly good cake removal at any knife advance. This was presumably caused by having exceeded the capacity of the rotating knives. This limitation upon drum speed may be counteracted by increasing the rotational speed of the doctoring knives.

* Product of Johns-Manville Corporation

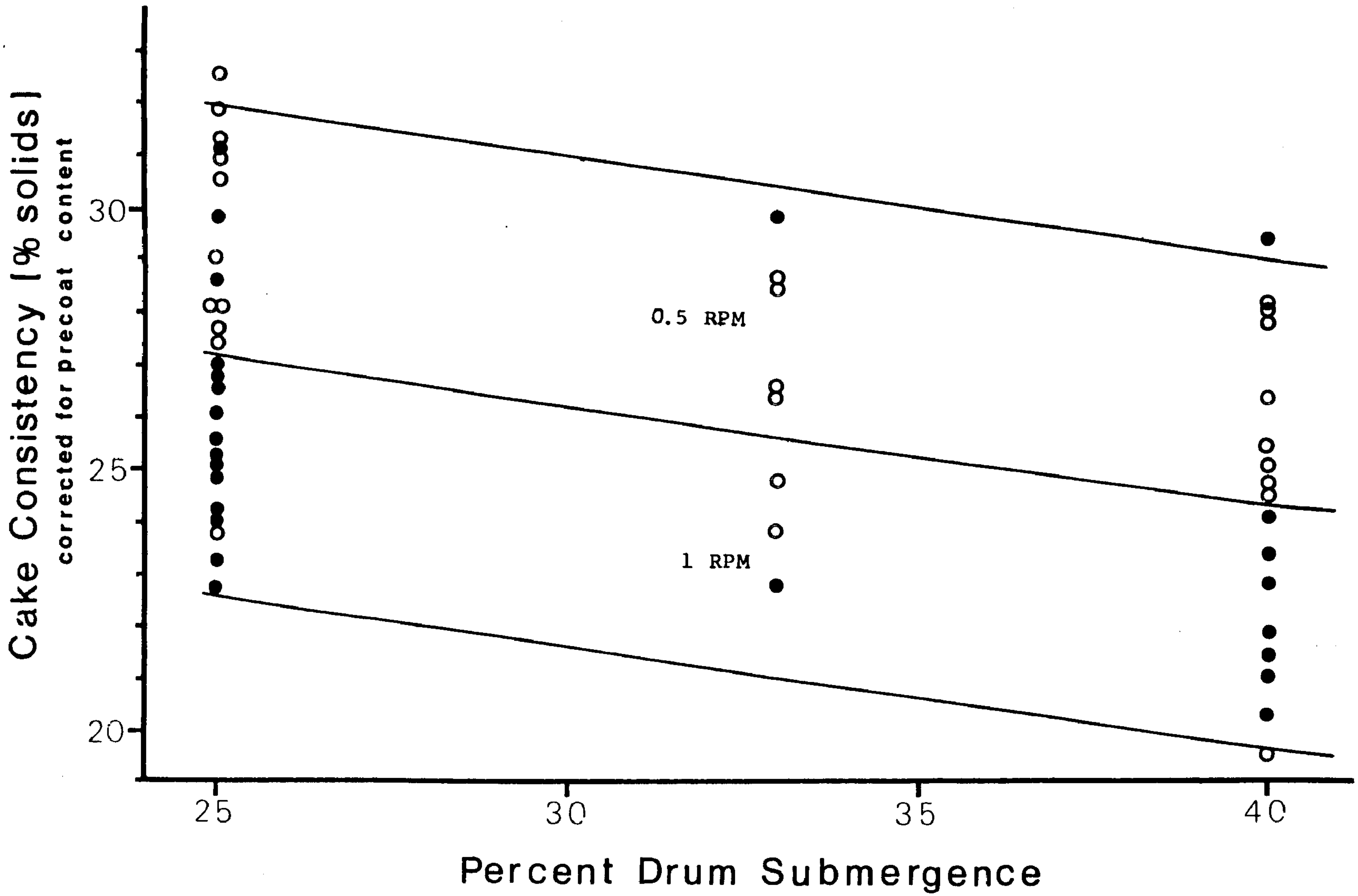


Figure 9. Precoat Vacuum Filter cake consistency.

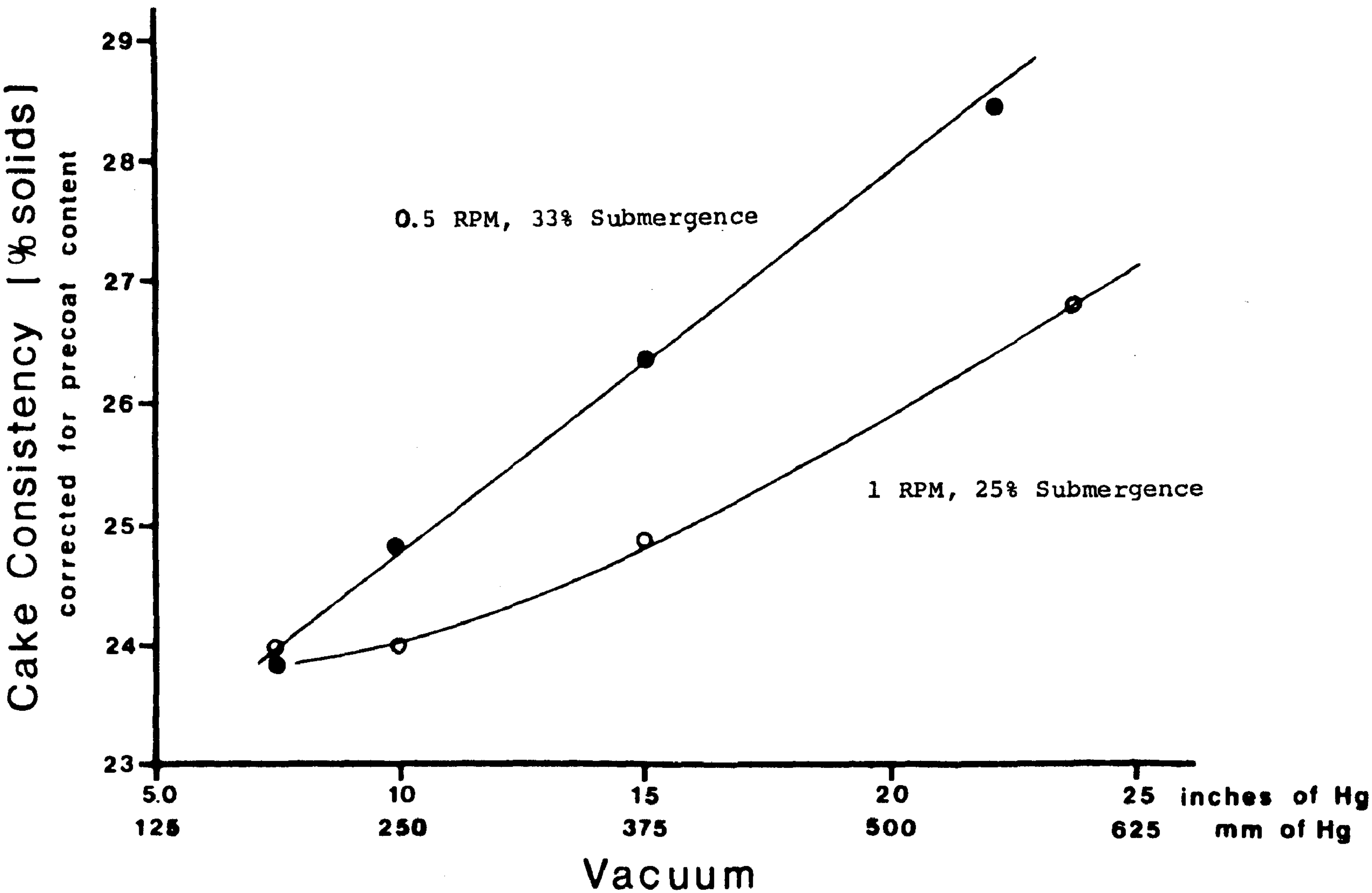


Figure 10. Effect of vacuum upon Precoat Vacuum Filter cake consistencies.

Alternative Precoat Materials

The investigation of precoat vacuum filtration included a preliminary assessment of the feasibility of utilizing materials other than diatomaceous earth for precoating. The pilot unit was operated with both flyash and lime mud ($CaCO_3$) precoats at different times in the test program. Table 9 summarizes the results of these tests. Flyash provided filtration rates, and cake consistencies quite comparable to diatomaceous earth.

As Table 10 indicates, the flyash solids particle size used in this pilot study was substantially larger than that of Celite 545, a relatively coarse grade of diatomaceous earth. In addition, as shown in Table 11 & 12(7), flyash solids size distribution and volatile content vary considerably from mill to mill, making it necessary to evaluate the potential for flyash precoat filtration on a case-by-case basis. Even at a particular mill, the day-to-day variations in flyash characteristics suggest that consideration might wisely be given to installation of size classification equipment to assure a consistent quality of precoat material. Utilization of flyash solids of such a size distribution that sludge solids freely penetrate the precoat body will necessitate a knife advance significantly greater than 1 mil (25.4 microns) per revolution to provide a filtering surface with acceptable permeability and may ultimately lead to media blinding. As shown in Table 9 and discussed in the solids recovery section, there was some indication that the flyash precoat body was penetrated by sludge solids in this study but not to a degree that solids loading rates were noticeably affected over the three-hour test period.

Lime mud ($CaCO_3$), as utilized in this program, resulted in reductions in sludge solids loading rates of roughly 50 percent. The lime mud precoat was prepared by reslurrying lime mud filter cake from the mill's recovery system to 5 to 10 percent consistency in fresh water. The application of a 1-in. (2.54 cm) thick lime mud precoat to the pilot filter from slurry prepared in this manner required 2 hours, suggesting that its filtration characteristics had been altered in preparation. Later laboratory investigations designed to determine the effect of precoat preparation upon lime mud filtration rates revealed that the specific resistance to filtration of lime mud from the lime mud filter vat (in the kraft recovery system), at 4.7 percent consistency and cooled to 104° F (40° C), was roughly one-half of the specific resistance measured when lime mud filter cake was reslurried in fresh water at the same temperature and consistency. It is not clear whether this phenomenon is attributable entirely to the nature of the mother liquor or to a combination of factors. In any case, the utilization of lime mud in the precoat filtration of waste treatment sludges will require a more complete understanding of lime mud filtration characteristics.

TABLE 9. EVALUATION OF ALTERNATIVE PRECOAT MATERIALS

Test No.		Precoat material	Feed consistency % solids	Drum speed RPM	% Drum submergence	Vacuum inches Hg	Filtrate rate $gal/ft^2/hr$	Solids loading rate $\#/ft^2/hr$	% Cake solids corrected for precoat content
1	a	D.E.*	Water	1.00	25	10	111.10	-	-
	b	Flyash	Water	1.00	25	10	2.16**	-	-
2	a	D.E.	0.97	0.50	25	21	4.47	0.36	28.1
	b	Flyash	1.78	0.50	25	21	5.72	0.85	26.0
3	a	D.E.	1.45	0.67	25	21	6.48	0.78	27.8
	b	Flyash	1.70	0.67	25	21	7.43	1.05	26.0
4	a	D.E.	1.07	0.67	40	21	8.72	0.78	23.4
	b	Flyash	1.01	0.67	40	21	8.40	0.71	24.0
5	a	D.E.	1.03	0.50	40	21	8.26	0.71	24.7
	b	Flyash	1.04	0.50	40	21	8.5	0.74	23.0
6	a	Lime mud	Water	1.00	25	10	3.20	-	-
	b	D.E.	Water	1.00	25	10	35.50	-	-
7	a	Lime mud	1.07	1.00	40	21	9.03	0.81	26.6
	b	D.E.	1.02	1.00	40	21	17.73	1.51	19.6
8	a	Lime mud	1.25	0.50	40	21	7.01	0.73	25.1
	b	D.E.	0.96	0.50	40	21	11.79	0.94	19.6
9	a	Lime mud	0.94	0.50	25	21	3.39	0.27	5.0
	b	D.E.	1.08	0.50	25	21	8.27	0.75	23.7
10	a	Lime mud	1.06	1.00	25	21	4.14	0.37	15.8
	b	D.E.	1.46	1.00	25	21	11.68	1.42	25.5

* D.E. is Celite 545 diatomaceous earth. Because this set of tests was not designed to optimize precoat utilization and because Celite 545 and flyash were expected to be relatively open precoating materials, knife advances as high as 3 ml/rev. (76 micron/rev.) were utilized.

** This fresh water filtrate rate was measured after the day's testing was completed. The low value may indicate that sludge solids have penetrated the precoat body.

TABLE 10. FLYASH - DIATOMACEOUS EARTH
SIZE DISTRIBUTION COMPARISON

Flyash		Celite 545		Hy-Flo	
Size range (microns)	% by weight	Size range (microns)	% by weight	Size range (microns)	% by weight
>250	12				
65-250	84				
40-65	3	40-60	16	40-60	4.0
20-40	1	20-40	31	20-40	11.0
10-20	0.2	10-20	41	10-20	30.0
<10	<0.01	8-10	6	8-10	12.5
		6-8	3	6-8	13.5
		<6	3	4-6	15.0
				2-4	10.0
				<2	4.0

TABLE 11 . FLYASH PARTICLE SIZE

Sample no.	Retained on mesh size (percent of total weight)												
	0*	1	2	3	4	5	6	7	8	9	10	11	12
20	4	0	0	12	12	0	21	14	0	2	0	2	0
50	-	10	18	52	23	2	71	14	1	50	19	59	14
60	8	5	4	5	18	2	3	4	1	5	5	4	4
80	11	13	9	12	25	4	3	12	2	25	30	14	13
100	8	19	13	9	10	21	2	14	3	12	18	8	13
120	-	13	13	4	6	6	0	12	5	5	11	5	13
150	13	-	-	-	-	-	-	-	-	-	-	-	-
200	-	27	21	4	5	23	0	15	17	1	11	5	20
<150	56	-	-	-	-	-	-	-	-	-	-	-	-
<200	-	13	22	2	1	42	0	15	71	0	6	3	23

TABLE 12. COMBUSTIBLE MATTER CONTENT OF FLYASH SAMPLES

Sample no.	% Combustible
0*	32
1	39
2	43
3	78
4	27
5	46
6	54
7	25
8	9
9	6
10	9
11	47
12	21

* Sample 0 used for precoating in this study. Other data is from reference 7.

Sludge Conditioning

The application of polymer conditioned sludge to the precoat vacuum filter resulted in approximately 50 percent increases in solids loading rates and decreases in cake consistencies of 7 to 8 percent solids. The degradation in cake consistencies, shown in Table 13, appeared to be caused by the thickness of the individual sludge flocs as they were deposited on the precoat. The deposited cake was uneven, consisting of raised individual flocs as opposed to a flat cake, the raised areas of the cake remaining wet throughout the filter cycle.

Overall Performance of Precoat Vacuum Filtration

The data generated during this pilot investigation suggested that the feasibility of precoat vacuum filtration would be greatly enhanced at this mill by thickening of the waste activated sludge from 0.7 to 0.8 percent consistency to 1.5 to 2.0 percent consistency. Depending upon the conditions of operation, such thickening could be expected to increase solids loading rates by from 80 to 300 percent. This mill's fresh waste activated sludge at 1.5 to 2 percent consistency could be dewatered at 1.5 to 1.75 $lb/ft^2/hr$ (7.33 to 8.55 $kg/m^2/hr$) generating cakes of 25 to 30 percent consistency after being corrected for precoat content (26 to 31 percent bulk consistency). Based upon a knife advance of 1 mil (25.4 microns) per revolution, diatomaceous earth consumption is calculated to be about 3 percent of the weight of sludge solids being dewatered. To achieve these performance levels, the filter would be operated at a vacuum of 20 to 22 in. (510 to 560 mm) of mercury, a form time of 48 seconds and a cycle time of 2 minutes.

Mills generating biological sludges with a specific resistance in the 200 x 10^7 to 1000 x 10^7 sec^2/gm range might realistically expect to attain solids loading rates of roughly 0.5 $lb/ft^2/hr$ (2.4 $kg/m^2/hr$) while generating cakes of 20 to 25 percent consistency after being corrected for precoat content (22 to 26 percent bulk consistency). Because a knife advance of 1 mil (25.4 microns) per revolution would again likely be required, the precoat consumption would be approximately 5 to 10 percent of the weight of sludge solids being dewatered.

In all instances, solids recoveries would be expected to exceed 99 percent.

TABLE 13. UTILIZATION OF POLYMER CONDITIONING IN PRECOAT VACUUM FILTRATION

Data set	Conditioning #polymer/ton	Feed Consistency # solids	Drum speed rpm	% Submergence	Knife advance mil/min	Loading rate $\#/ft^2/hr$	Cake consistency % solids
1	0	1.53	0.50	25	0.5	1.13	31.5
1	5	2.78	0.50	25	0.5	2.24	23.1
2	0	0.94	0.50	40	0.5	0.89	26.4
2	5	1.24	0.50	40	0.5	1.73	18.6
3	0	1.62	0.67	25	1.0	1.05	30.6
3	5	1.21	0.67	25	1.0	1.31	23.1
4	0	0.94	0.67	40	1.0	0.93	25.2
4	5	1.45	0.67	40	1.0	2.23	18.5

SECTION 7

TAIT-ANDRITZ SLUDGE DEWATERING MACHINE (SDM) FILTER BELT PRESS INVESTIGATION

PROCESS DESCRIPTION

The Tait-Andritz SDM filter belt press, shown schematically in Figure 11, dewaters sludges through a combination of gravity, pressure and shear forces. A free-draining sludge is applied to the belt and allowed to undergo gravity drainage over the first section of the press. The sludge is then met by a second belt, and conveyed through a wedge section where the two belts gradually converge. In this section, dewatering is achieved by both the pressure applied by the belts and by table rolls behind both top and bottom belts. The two belts converge at the beginning of the "S" press section where they travel an "S" shaped path around several rollers, as a result of which the belts are forced to move at slightly different speeds imparting shear to the sludge cake. The pressure exerted upon the sludge is controlled through adjusting the tension on the belts. The cake is discharged by doctor blades on both top and bottom belts. Both belts are shower cleaned while returning to the head of the machine. The SDM also can be equipped with a "P" press section consisting of a series of nips for dewatering fibrous sludges. The variables of importance to SDM performance are listed in Table 14 .

DESCRIPTION OF TAIT-ANDRITZ SDM

The unit used in this study was the smallest production model offered by Tait-Andritz. The belt was 22 in. (56 cm) wide, the entire unit being about 12 feet (3.7 m) long. Belt tension was controlled pneumatically, allowable air pressure ranging from 0 to 3 atmospheres (bars). The two belt materials evaluated were 60- and 30-mesh plain weave polyester fabrics.

OPERATION OF THE TAIT-ANDRITZ SDM

In the conduct of the pilot study the feed rate was first adjusted to the desired value. It was then necessary to determine the belt speed and polymer dosage combinations that

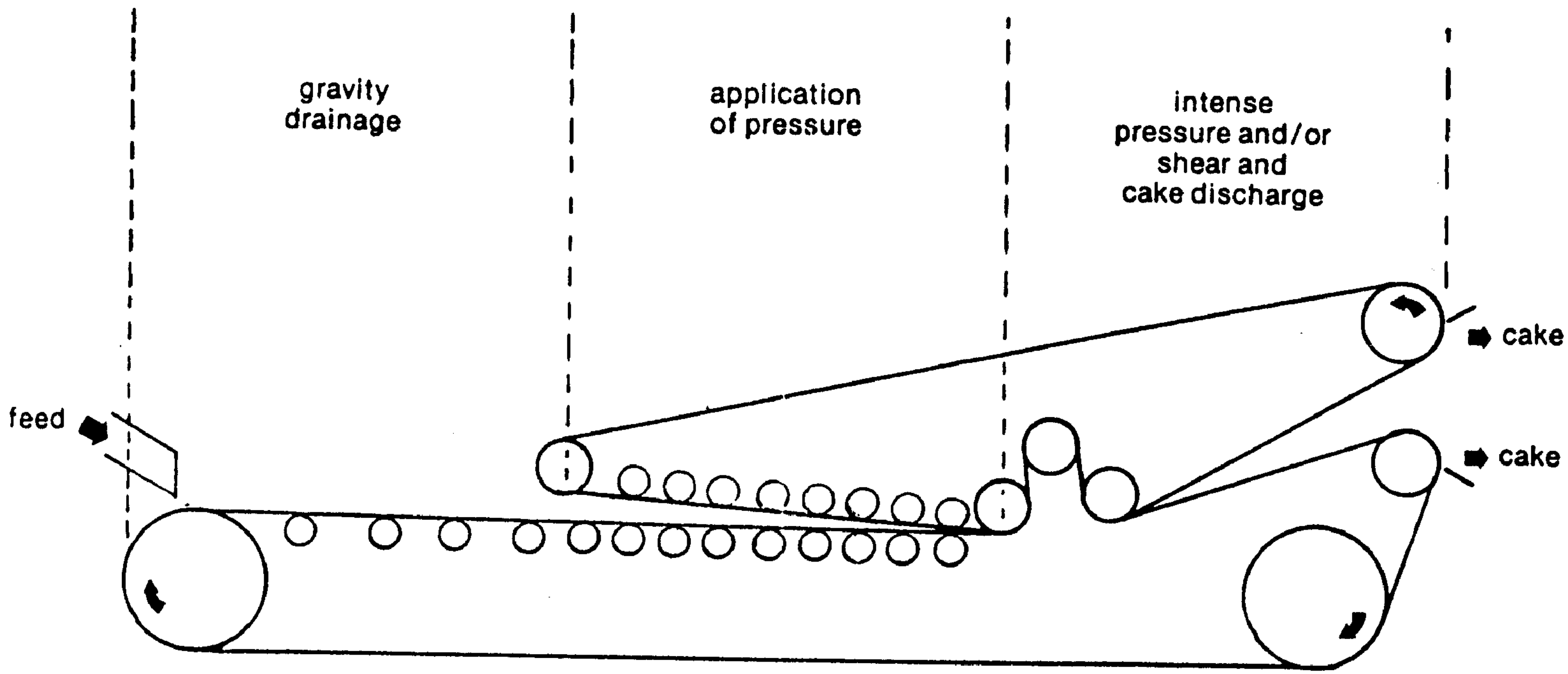


Figure 11. Tait-Andritz SDM.

TABLE 14. TAIT ANDRITZ SDM OPERATING VARIABLES

Independent variables
1. Belt characteristics
2. Type of press section ("S" or "P")
3. Belt tension and/or nip pressure
4. Belt speed
5. Feed rate
6. Feed consistency
7. Sludge conditioning
8. Nature of the sludge solids
Dependent variables
1. Cake consistency
2. Solids recovery

produced stable operation as manifested by a nonincreasing liquid level in the gravity drainage section. After an acceptable set of operating conditions had been established, a 5- to 10-minute stabilization period was allowed before collecting samples of unconditioned feed, conditioned feed, gravity drainage filtrate, press section filtrate, belt shower water and cake. The flow rates of the two filtrate streams and shower water stream were measured and the cake generation rate was documented. The evaluation was then continued by varying belt speed and belt tension at the feed rate and polymer dosage of interest, samples and flow rates being taken at each set of conditions. Then either the polymer dosage and/or the feed rate were adjusted to different levels.

TAIT-ANDRITZ SDM RESULTS

Capacity

The rate at which the 20-inch (500 mm) unit could dewater fresh waste activated sludge at 0.8 percent consistency was determined by the rate at which drainage occurred prior to the sludge's entering of the press section, and thus was determined by polymer dosage. The polymer requirements for different feed rates of fresh waste activated sludge (unconditioned specific resistance less than 150 x 10 sec/gm) are shown in Table 15. The table suggests that the sludge had to be conditioned to a specific resistance of 30 x 10^7 sec^2/gm or less to provide acceptable throughputs. The belt was run at the slowest speed possible while maintaining an acceptable (not overflowing)

TABLE 15. TAIT-ANDRITZ SLUDGE DEWATERING MACHINE (SDM)

PERFORMANCE SUMMARY ON FRESH WASTE ACTIVATED SLUDGE:
FEED CONSISTENCY OF 0.8 PERCENT SOLIDS

Feed rate per unit belt width		Polymer requirement Betz 1260		Belt speed		Specific resistance of conditioned sludge
g/min/ft	(l/min/m)	#/tn	(mg/gm)	ft/min	(cm/min)	$\times 10^7$ sec^2/gm
4.7- 5.6	(60- 70)	5-10	(2.5-5)	6.6-8.2	(200-250)	15-30
10.3-11.2	(130-140)	5-10	(2.5-5)	6.6-8.2	(200-250)	15-30
12.5-12.0	(160-170)	5-10	(2.5-5)	8.2-9.8	(250-300)	15-30
18.1-19.4	(220-240)	8-12	(4.0-6)	16.4	(500)	10-20
23.1-31.7	(290-390)	10-15	(5-7.5)	18.0-21.3	(550-650)	<15

liquid level in the gravity drainage and wedge sections. The maximum belt speed corresponded to the occurrence of solids extrusion at the end of the wedge section due to the cake being too wet for pressing. Maximum belt speeds were seldom encountered at low feed rates, 5 to 7 gpm (20 to 25 liters/min), whereas at the other extreme, 40 gpm (150 liters/min), there was very little difference between maximum and minimum belt speeds. Because of the strong dependence of machine capacity upon liquid drainage rate, it is anticipated that prethickening of sludge feed would allow substantial improvements in solids handling capacity. The data were insufficient to allow a determination of polymer requirements on sludge with an unconditioned specific resistance in excess of 150 x 10^7 sec^2/gm.

Cake Consistencies

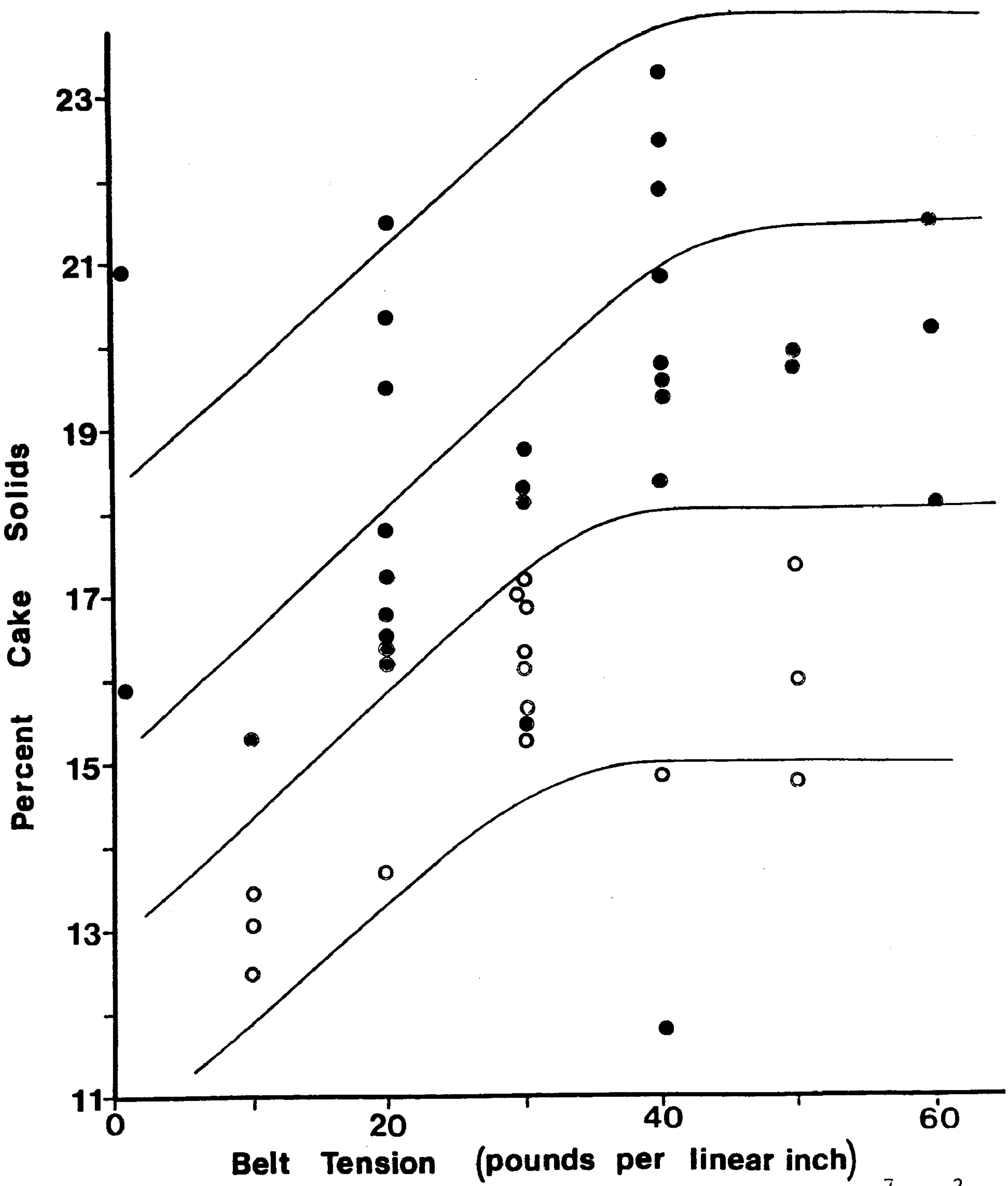
The cake consistencies attained by the SDM were determined by the belt tension utilized and the nature of the sludge solids. Figure 12 demonstrates that the highest cake consistencies were associated with unconditioned fresh sludges, and the lowest with conditioned, physically degraded (sheared or aged) sludges having unconditioned specific resistances in excess of 150 x 10^7 sec^2/gm. Conditioned fresh waste activated sludge cake consistencies fell between these other two. Belt tensions in excess of 40 pounds per linear inch were not beneficial. Varying belt speed within the range of 4.3 to 21 ft/min (130 to 650 cm/min) did not generally have a significant impact upon cake consistency. There was some indication that belt mesh also impacted upon cake consistencies, with a 30-mesh belt providing drier cakes than a 60-mesh belt.

Solids Recoveries

There were three different types of solids lost from the SDM: those which either (a) passed through the belt in the drainage sections, (b) overflowed the drainage sections due to insufficient conditioning and/or belt speed, or (c) washed from the media in the shower section. The shower water solids were particularly evident at high belt tensions, accounting for an average of 54 percent of the unrecovered solids at belt tensions of 50 pounds per linear inch, and an average of 38 percent of the unrecovered solids at belt tensions less than 50 pounds per inch. The SDM solids recoveries were consistantly in excess of 85 percent with 90 percent recovery being representative during periods of operaiton at likely operating conditions. Insufficient sludge conditioning resulted in limited capacity before solids recovery was severely affected.

Conditioning Requirements

As previously demonstrated in Table 15 , conditioning



● Unconditioned Sludge, Specific Resistance <150 x 10^7 sec^2/gm

◉ Conditioned Sludge, Unconditioned Specific Resistance <150 x 10^7 sec^2/gm

○ Conditioned Sludge, Unconditioned Specific Resistance >150 x 10^7 sec^2/gm

Figure 12. Tait-Andritz SDM cake consistencies.

requirements were primarily dictated by feed rate. Higher feed rates required more conditioning.

Overall Performance of the Tait-Andritz SDM

The unit was capable of dewatering up to 45 gpm (165 liters/min) with conditioning requirements of up to 15 pounds Betz 1260/ton (7.5 mg/gm). A feed rate of 20 to 25 gpm (76 to 95 liters/min) conditioned with 5 to 10 pounds of Betz 1260/ton (2.5 to 5 mg/gm) was found to be the maximum throughput at which consistent results were assured. At these conditions, fresh waste activated sludge was concentrated from a feed consistency of 0.8 percent solids to a cake consistency of 17 to 20 percent solids at 90 percent recovery.

SECTION 8

PERMUTIT DUAL CELL GRAVITY FILTER MULTIPLE ROLL PRESS INVESTIGATION

PROCESS DESCRIPTION

The Permutit DCG-MRP is a two-stage moving belt press dewatering system consisting of a dual cell gravity filter (DCG) and a multiple roll press (MRP). The DCG, shown schematically in Figure 13, consists of two parallel, cylindrical, media-covered chambers, connected by virtue of being covered by the same piece of media. A well-conditioned sludge is fed to the first chamber where the initial dewatering occurs, removing a majority of the free water. As this initial dewatering takes place, sludge solids are deposited upon the moving media and transported over the raised interchamber barrier into the second cell. In this second cell the sludge solids agglomerate into a flexible "plug" which continuously "rolls" as the drum rotates. This rolling and kneading action releases additional water. The plug grows in size until segments of it at the ends of the cell overflow retaining rims and fall onto a conveyor belt to be transported to the MRP.

The MRP, also shown in Figure 13, is a twin belt, multiple roll press containing a series of opposing and alternating rollers between which the DCG cake passes, releasing additional water. Both roller pressure and belt speed are variable. The cake is discharged from the media by doctor blades.

The variables of importance to DCG and MRP operation are listed In Table 16.

DESCRIPTION OF DUAL CELL GRAVITY FILTER AND MULTIPLE ROLL PRESS PILOT UNIT

Dual Cell Gravity Filter (DCG)

The unit used in this study was the smallest production model of the DCG, the DCG 100. The two cells each measured 20.5 in. (52 cm) in diameter by 23.5 in. (60 cm) long and were 27 in. (69 cm) from center to center. Two different media were

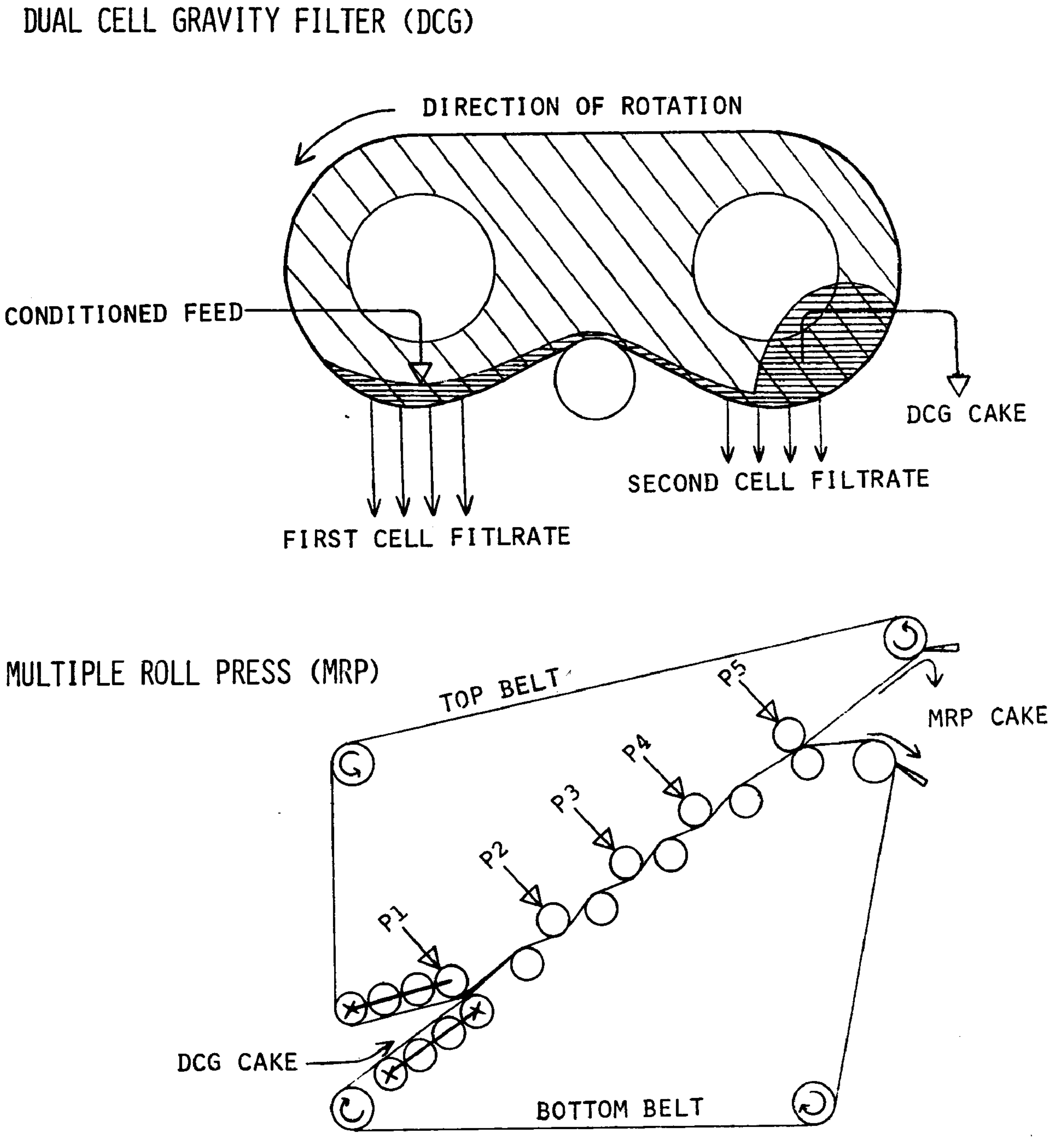


Figure 13. Permutit DCG-MRP.

TABLE 16. DUAL CELL GRAVITY FILTER (DCG) AND MULTIPLE ROLL PRESS (MRP) PROCESS VARIABLES

Dual Cell Gravity Filter (DCG)

Independent variables

1. Screen characteristics
2. Screen speed
3. Feed rate
4. Feed consistency
5. Nature of the feed solids
6. Sludge conditioning

Dependent variables

1. Cake consistency
2. Solids recovery

Multiple Roll Press (MRP)

Independent variables

1. Belt characteristics
2. Belt speed
3. Roller pressure
4. Feed rate
5. Feed consistency
6. Nature of the feed solids
7. Sludge conditioning

Dependent variables

1. Cake consistency
2. Solids recovery

available, one being 100 mesh, the other 40 mesh. The unit was equipped with variable speed belt drive. The filter media was washed continuously by showers located above the first cell.

Multiple Roll Press (MRP)

The MRP unit tested was a production unit normally capable of dewatering the output from three DCG's 100 units. The belts were 36 in. (91 cm) wide and the unit measured 5 feet (1.5 m) from the feed point to the doctor blades. The belts first passed through a series of rollers designed to apply a gradually increasing pressure to the sludge cake. This first set of rollers was mounted together, the upstream end of the set being fixed at a pivot point and the downstream end floating at an adjustable pressure. The belt then traveled a serpentine route through three sets of alternating offset rollers, then between a pair of opposing rollers. The pressures applied to all sets of offset and opposing rollers were adjustable. Cake removal was accomplished with doctor blades on both top and bottom belts. Belt showers were applied to both belts during their return to the front end of the machine. The MRP belts were laminations of 100-mesh surface media and coarser support material held together with several ribs of bonding material running parallel to the direction of belt travel.

OPERATION OF THE DUAL CELL GRAVITY FILTER (DCG) AND MULTIPLE ROLL PRESS (MRP)

Test runs were initiated by adjusting feed rate, conditioning levels, DCG belt speed, MRP belt speed, and the MRP roller pressure profile to the desired values. The DCG was then observed for approximately 5 minutes to determine whether the set of operating conditions was a stable set. On occasions when the sludge conditioning levels were insufficient for the feed rate being applied, the first DCG cell liquid level would rise and, when given the opportunity, overflow from the cell. If it was clear that this was about to occur, the set of circumstances was recorded and the conditioning was increased until sufficient to permit free drainage. Due to the relative capacities of the DCG and MRP, the latter was never fully loaded. After a determination that the operating conditions were acceptable, samples were collected of unconditioned feed, conditioned feed, DCG filtrate, DCG cake, MRP filtrate and MRP cake. The flow rates of the DCG feed, DCG filtrate and MRP filtrate were recorded. Specific resistance tests were conducted on all unconditioned and conditioned feed samples to detect changes in the nature of the solids and to assess the effectiveness of the applied sludge conditioning.

DUAL CELL GRAVITY FILTER (DCG) AND MULTIPLE ROLL PRESS (MRP) TEST RESULTS

Loading Rates

Capacities were determined by the rate at which drainage occurred in the first DCG cell, which reflected the amount of conditioning agent utilized. Table 17 shows the amount of conditioner required to dewater various feed rates of fresh waste activated sludge (unconditioned specific resistance of 30 to 250 x 10^7 sec^2/gm) and degraded waste activated sludge (unconditioned specific resistance of 110 to 350 x 10^7 sec^2/gm). Higher

TABLE 17. DCG 100 PERFORMANCE SUMMARY

Feed rate gpm	(l/min)	#/hr	(kg/hr)	Polymer requirement Betz 1260 #/tn	(mg/gm)	Conditioned specific resistance required x 10^7 sec^2/gm
FRESH WASTE ACTIVATED SLUDGE AT 0.8 PERCENT CONSISTENCY						
5	(19)	20	(9)	5	(2.5)	30-40
10	(38)	40	(18)	5-10	(2.5-5)	20-40
15	(57)	60	(27)	5-10	(2.5-5)	10-40
HIGHLY SHEARED OR AGED WASTE ACTIVATED SLUDGE AT 0.8 PERCENT CONSISTENCY						
5	(19)	20	(9)	5	(2.5)	30-40
10	(38)	40	(18)	8-10	(4-5)	20-40
15	(57)	60	(27)	10-15	(5-7.5)	10-20
18	(68)	72	(33)	20	(10)	5

feed rates required larger polymer dosages. Table 16 also indicates that the difference between dewatering fresh and degraded sludge solids was the amount of conditioning required to bring them into the range of specific resistance necessary for handling a given feed rate. Degraded sludges required more conditioning than fresh sludges, especially at higher feed rates.

Although the most important factor in maintaining acceptable loading rates was the level of conditioning utilized, DCG belt speed was also a factor. In one instance where the sludge was not quite sufficiently conditioned, an increase in belt speed of from 57 to 70 in./min (145 to 180 cm/min) was necessary to prevent the first cell from overflowing. It is likely that the same effect could have been attained by slightly decreasing the feed rate or increasing the amount of conditioning. One set of tests conducted with a thickened feed at 1.2 percent consistency (as opposed to an unthickened consistency of 0.8 percent) did not demonstrate any significant improvements in performance attributable to the increase of 0.4 percent in feed concentration. No significant increase in capacity was associated with the use of a 40-mesh over a 100-mesh media fabric.

The relative differences in capacity between the DCG and MRP did not allow a determination of the loading rate limitations of the MRP. However, based upon observation of the amount of unused belt at the discharge end of the MRP, it was estimated that the unit could dewater the output from three Dual Cell Gravity filters.

Consistencies of Dual Cell Gravity Filter (DCG) Cakes

The most striking aspect of the cake consistencies attained on the DCG was their uniformity. Thirty-six of the forty-six cakes generated on polymer conditioned sludges had consistencies of 7.8 to 9.8 percent solids. The total range of cake consistencies was 6.8 to 9.8 percent solids. Variations in conditioning, specific resistance, belt speed, belt mesh and feed consistency did not have a significant impact upon DCG cake consistencies.

Consistency of Multiple Roll Press (MRP) Cakes

The cake consistencies attained by the MRP were determined by the roller pressure at the final nip which was the highest pressure applied. Figure 14 demonstrates this effect as well as the extent of the loss of consistency attributable to high belt speeds. Analysis of the data revealed that the scatter in cake consistency at a constant final nip pressure was not due to differences in the total pressure applied by the other rollers. The MRP was not loaded heavily enough in this study to encounter solids extrusion problems attributable to excessive initial pressures at the first MRP rollers.

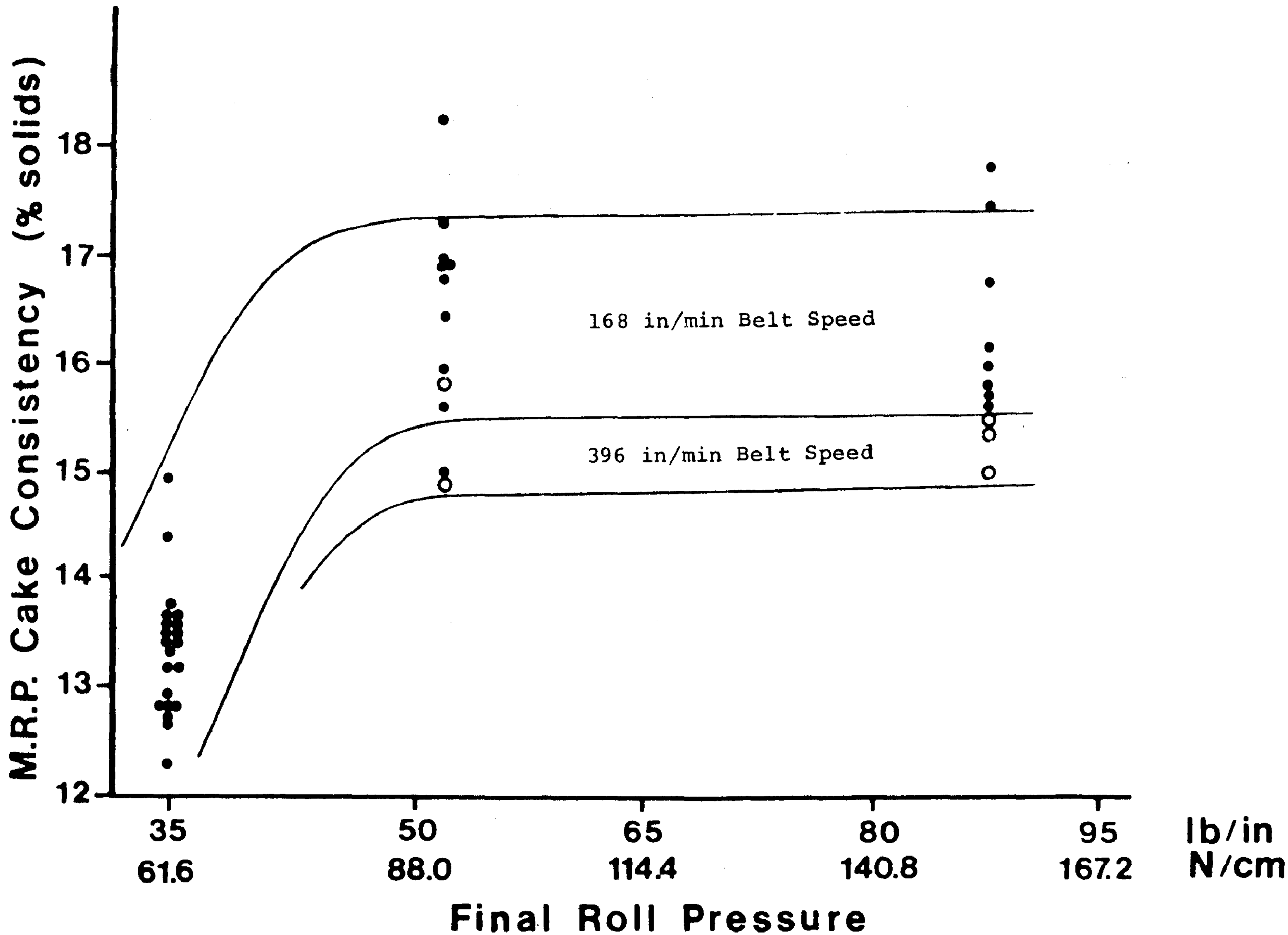


Figure 14. MRP cake consistencies.

Solids Recoveries Attained by the Dual Cell Gravity Filter (DCG)

The lowest solids recovery attained by the DCG when operating under stable (no overflow) conditions was 97.2 percent. Twenty-five of the thirty-five recorded values for DCG solids recovery were 99 percent or higher. In essence, the conditioning requirements for maintaining acceptable throughputs were such that solids recoveries were always high. The solids recoveries associated with the use of the 40-mesh DCG screen were not significantly different from those attained with the 100-mesh screen.

Solids Recoveries Attained by the Multiple Roll Press

The belts utilized on the MRP were laminated and bonded together with plastic ribs. Because these ribs were slightly raised from the media surface, difficulties were encountered in accomplishing complete cake discharge. The solids not removed by the doctor blades ultimately contributed to the solids in the filtrate causing decreased solids recovery. In spite of this difficulty, 31 of the 41 tests for MRP solids capture showed values exceeding 90 percent, with 21 of the 41 exceeding 95 percent.

Conditioning Requirements

As was shown in the discussion of DCG capacity, the conditioning requirements were related to the feed rate and the nature of the solids, higher feed rates and less filterable sludges requiring more conditioning. In all instances, conditioning was necessary to promote drainage rather than to capture solids. In addition, all cakes generated by the DCG were sufficiently conditioned to render them pressable in the MRP.

Overall Dual Cell Gravity Filter (DCG) and Multiple Roll Press (MRP) Performance

The DCG was capable of dewatering up to about 20 gpm of waste activated sludge at 0.8 percent consistency, generating cakes at 9 percent consistency, and requiring sludge conditioning at from 5 to 20 lb of polymer/ton of sludge solids depending upon the feed rate and the nature of the solids. DCG solids recoveries were generally in excess of 99 percent. The MRP, operating with a great deal of excess capacity, dewatered the DCG cake from 9 percent to a cake consistency of 16 percent, attaining solids recoveries generally in excess of 90 percent.

SECTION 9

CAPILLARY SUCTION SLUDGE DEWATERING DEVICE INVESTIGATION

PROCESS DESCRIPTION

Biological and other difficult-to-dewater sludges require partial dewatering prior to being pressed. This is usually accomplished through gravity or vacuum assisted drainage. In the case of most filter belt presses, a relatively free-draining and therefore, well-conditioned sludge is required to provide acceptable throughputs. As an alternative to gravity drainage, the capillary suction sludge dewatering device, the Squeegee*, shown schematically in Figure 15, utilizes capillary suction to provide a cake suitable for pressing, potentially allowing a reduction in the amount of conditioning required. The other unique feature of this unit is the cake discharge mechanism. Whereas other filter belt presses utilize doctor blades to remove the sludge cake from the filter belt, the Squeegee transfers the cake to a combination pressure and pickup roll which is in turn doctored. The main components of the system are (a) a continuous, traveling screen to support the sludge above the capillary belt, (b) a porous belt which provides dewatering through the screen by capillary suction, (c) a combination pressure pickup roll which provides additional water removal and removes the resulting cake from the screen, (d) a doctor blade which scrapes the cake from the pickup roll, and (e) a set of opposing rollers to force water from the porous belt on its return trip to the head of the machine. The variables of consequence to Squeegee performance are listed in Table 18.

SQUEEGEE PILOT UNIT DESCRIPTION

The capillary suction belt press used in this study was a prototype Squeegee unit designed and built by Westinghouse Corporation under EPA contract (8). The belt measured 51.5 in. (130 cm) from the front lip of the feed tray to the center line of the pickup roller and was 11.8 in. (30 cm) wide. The pickup roller was 9 in. (23 cm) in diameter. The force exerted through the pickup roll was applied by a set of springs and adjusted by varying the spring lengths. Conditioning chemicals were injected into the feed line between the sludge pump and the

* Designed and built by Westinghouse Corporation

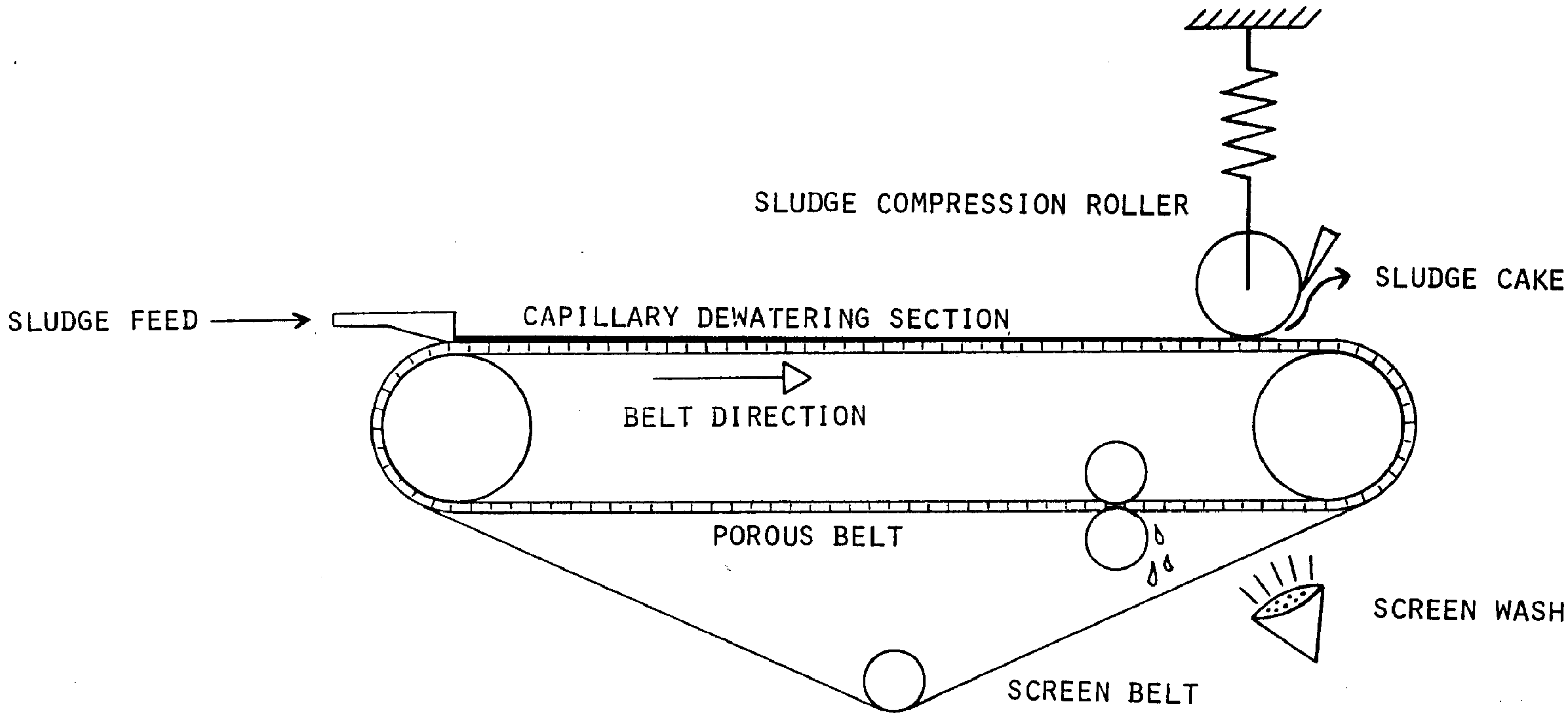


Figure 15. Schematic of Squeegee.

TABLE 18. SQUEEGEE PROCESS VARIABLES

Independent variables
1. Belt and screen characteristics
2. Belt speed
3. Feed position
4. Pickup roll diameter
5. Pickup roll pressure
6. Feed rate
7. Feed consistency
8. Sludge conditioning
9. Nature of the sludge solids
Dependent variables
1. Cake consistency
2. Solids recovery

head box of the machine, mixing being provided by turbulence in the feed line.

Screen showers were utilized to wash entrapped solids from the 80-mesh, plain weave screen.

OPERATION OF THE SQUEEGEE

The test runs were initiated by adjusting feed rate, conditioning rate, pickup roll pressure and belt speed to the desired values. The machine was then allowed to stabilize for 5 to 15 minutes, depending upon conditions, before samples were taken. The samples collected were unconditioned feed, conditioned feed, cake samples from several points along the belt, final cake, and filtrate. Flow rates of feed, filtrate and, on occasion, screen shower water were determined. All unconditioned and conditioned feed samples underwent specific resistance determinations to monitor changes in sludge dewaterability and to assess the effectiveness of the conditioning being applied.

SQUEEGEE RESULTS

Squeegee Capacity

The single most important factor in attaining acceptable performance from the Squeegee was the maintenance of complete cake removal by the pickup roll. Any operating conditions which caused incomplete removal quickly resulted in severe degradation of filtrate quality and screen blinding. These incidents tended to be related to excursions from certain ranges of belt speeds. The range of acceptable belt speeds varied with the feed rate, as indicated in Figure 16. As demonstrated in the Figure, unacceptable performance was associated with (a) excessive belt loadings at low belt speeds causing poor cake removal, (b) excessive belt speeds, causing poor cake removal, and (c) excessive belt speeds causing incomplete belt coverage at low feed rates. The fact that allowable belt speeds were more closely related to hydraulic than solids feed rates suggests that the unit was hydraulically limited, and might benefit from sludge prethickening. Figures 17 through 20 show that maintaining stable cake consistencies and/or solids recoveries while dewatering conditioned fresh waste activated sludge required a belt speed of less than about 3.4 in./sec (8.5 cm/sec). This being the maximum belt speed, Figure 16 indicates a maximum feed rate to be 2.6 gpm of adequately conditioned sludge per foot of belt width (32 liters/min/m). The maximum feed rate possible with unconditioned fresh waste activated sludge (specific resistance between 35 and 100 x 10^7 sec^2/gm) was found to be 1.3 gpm/ft of belt width (16 liters/min/m of belt width).

In general, to assure acceptable performance, the Squeegee required a ferric chloride conditioned sludge with a conditioned specific resistance of less than 30 x 10^7 sec^2/gm. Sludges which had been intentionally physically degraded by subjecting them to severe shear in a centrifugal pump did not readily condition to 30 x 10^7 sec^2/gm and as a result, could only be handled at rates less than 2.6 gpm/ft of belt width (32 liters/min/m). The amount of ferric chloride required to compensate for the physical degradation and permit loading rates comparable to fresh sludge was not determined.

In general, the capacity of the machine could be increased either by widening the belt, lengthening the belt (consistent with belt speed limitations) or both.

Cake Consistencies

The cake consistencies attained by the Squeegee on fresh conditioned sludge were found to be functions of the roll pressure and belt speed. Figures 17 through 19 constructed from all data generated with fresh, adequately ferric chloride conditioned, waste activated sludge (conditioned to a specific resistance of less than 30 x 10^7 sec^2/gm) demonstrate that consistent cake quality was attained at belt speeds of less than 3.0 to 3.5 in./sec (7.6 to 8.9 cm/sec). Unconditioned or poorly conditioned sludges (conditioned specific resistance of greater than 30 x 10^7 sec^2/gm) produced cakes at consistencies 3 to 5 percent lower than those generated from well conditioned sludges.

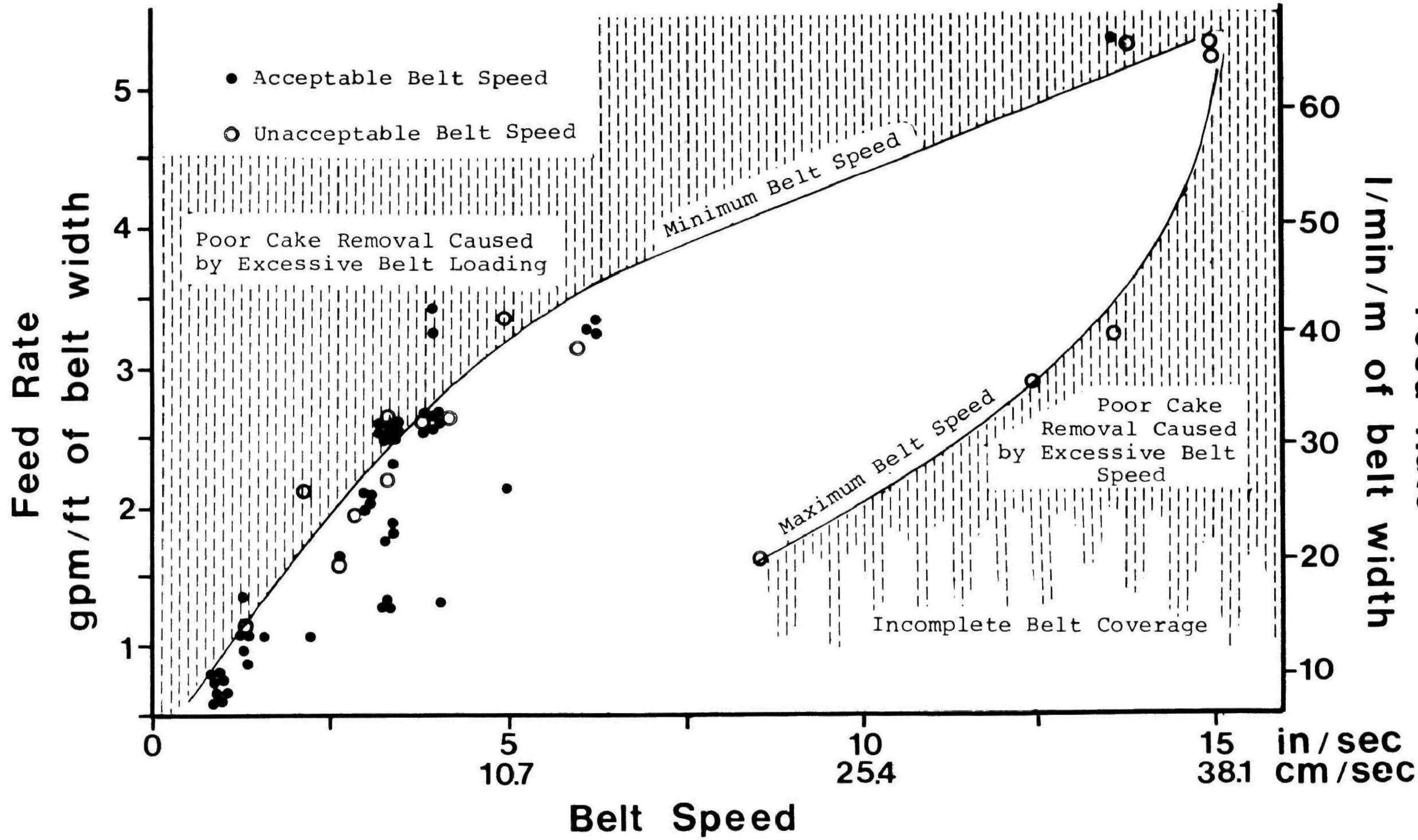


Figure 16. Squeegee belt speed requirements.

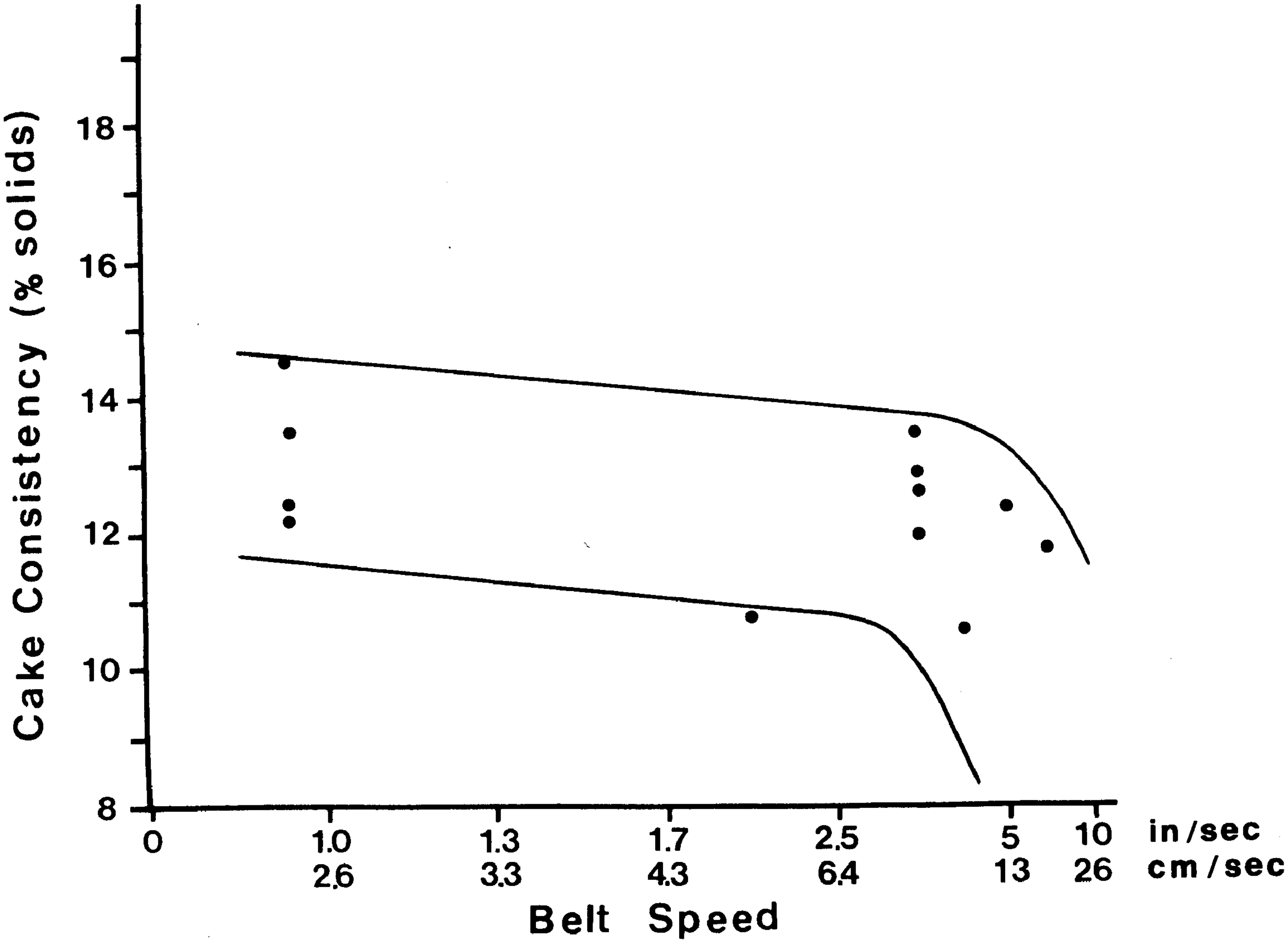


Figure 17. Squeegee cake consistencies at 0.6-1.7 #/linear inch of roller pressure (1-3 N/cm).

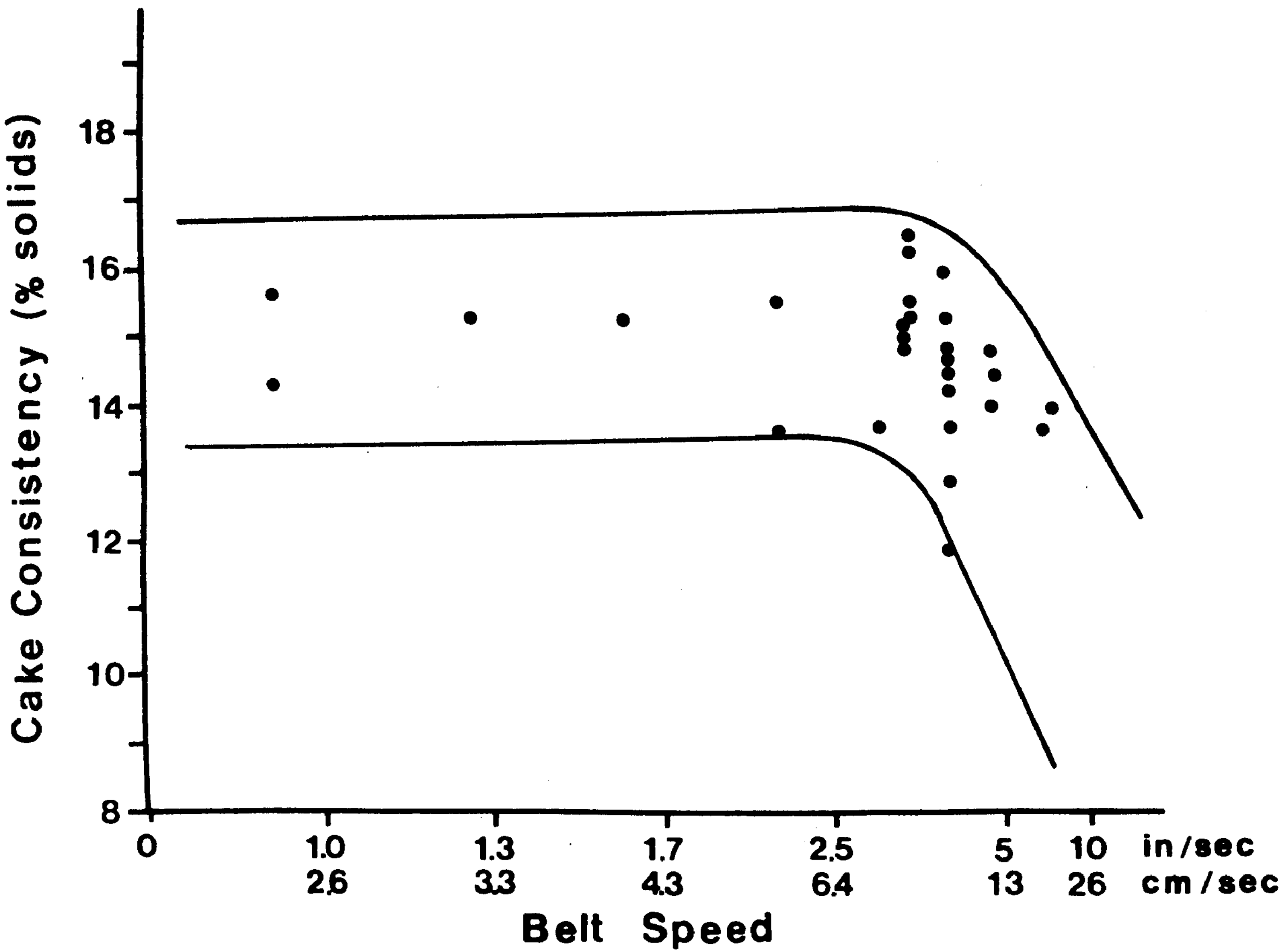


Figure 18. Squeegee cake consistencies at 3.9-4.5 #/linear inch of roller pressure (6.8-7.8 N/cm).

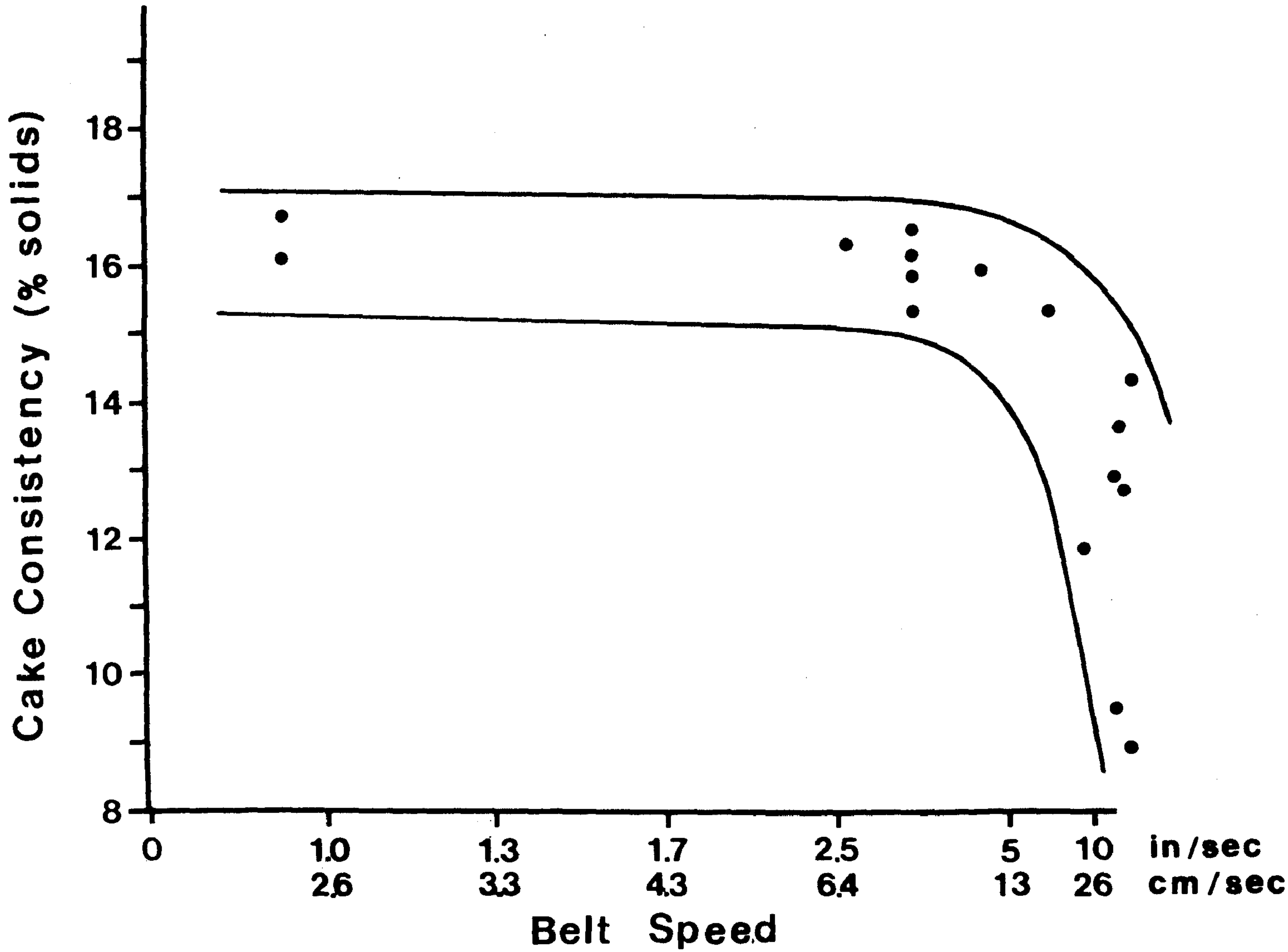


Figure 19. Squeegee cake consistencies at 6.1-7.3 #/linear inch of roller pressure (11-13 N/cm).

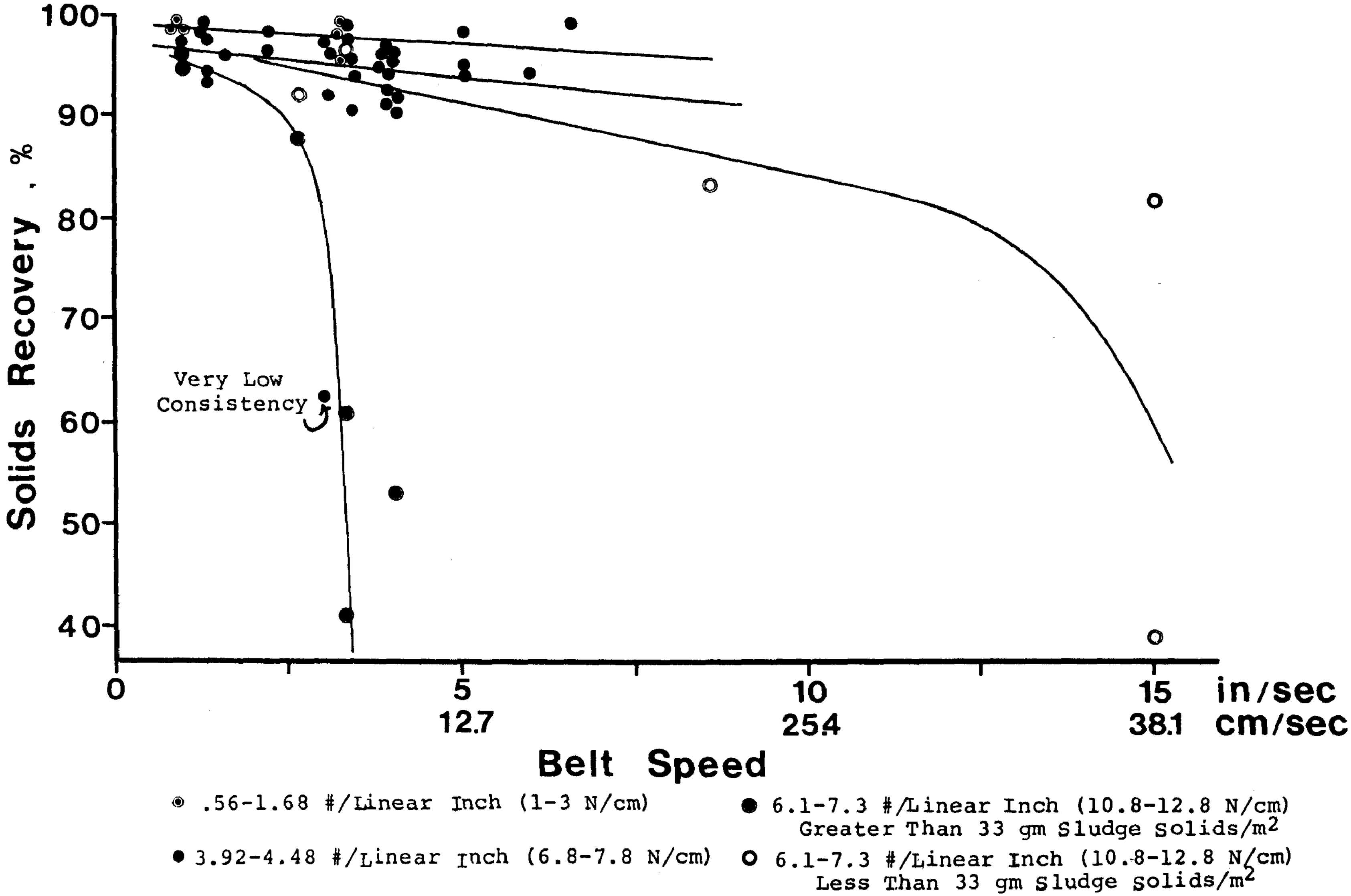


Figure 20. Squeegee solids recovery.

Solids Recovery

Solids recovery on sludges conditioned to a specific resistance of 30 x 10^7 sec^2/gm or less was generally maintained above 90 percent. On fresh waste activated sludge, this degree of conditioning was attained with additions of 4 to 7 percent ferric chloride. At high belt speeds and high roll pressures, cake removal became imparied resulting in reduced capture efficiencies, the onset of these difficulties occurring at lower belt speeds when belt loadings exceeded 33 gm of sludge solids per square meter as shown in Figure 20.

Conditioning Requirements

Because the cake removal properties were very closely related to the type and amount of sludge conditioning, these were possibly the most crucial controllable variables investigated. Attempts were made at dewatering unconditioned sludges and those sludges conditioned with either Betz 1260 (a cationic polyelectrolyte used successfully on the other filter belt presses), Percol 140 (a cationic polyelectrolyte), Hercufloc 859 (a cationic polyelectrolyte), and ferric chloride. Quite simply, ferric chloride was the only conditioning chemical applied which resulted in consistent performance at reasonable hydraulic loadings (greater than 25 liters/m^2). Dewatering without sludge conditioning was only possible at very low loadings and machine capacities, 1.3 gpm/ft of belt width (16 liters/min/m) being the maximum feed rate attainable. Polymer conditioning produced large, discrete floc that were impossible to distribute evenly across media. The tops of these floc remained wet virtually regardless of belt speed, making complete cake removal very difficult. Ferric chloride was the only conditioning technique evaluated that promoted rapid drainage while allowing the formation of a thin, even layer of sludge solids upon the screen which could be consistantly removed by the pickup roll. Ferric chloride was required at 4 to 7 percent of sludge solids for feed rates of 1.3 to 3.2 gpm/ft of belt width (16 to 40 liters/min/m) and 7 to 10 percent for feed rates between 3.2 to 5.2 gpm/ft of belt width (40 and 65 liters/min/m). However, above 2.5 gpm/ft poor cake removal or decreased cake consistencies were often encountered.

Squeegee Overall Performance

The unit used in this study was capable of dewatering 2.6 gpm/ft of belt width (32 liters/min/m) of fresh waste activated sludge from a feed consistency of 0.6 to 1 percent to a cake consistency of 14 to 17 percent solids. The data provided indicates that at feed consistencies of 0.6 to 1 percent the unit was hydraulically limited, suggesting sludge prethickening as one alternative for increasing machine capacity. Ferric chloride conditioning was required at 4 to 7 percent of sludge solids to promote rapid drainage and assure good cake discharge characteristics. Solids recovery was typically in excess of 90 percent. The fact that cake discharge was of such overriding importance to the operation suggests that improvements in media characteristics would be a possible means of substantially improving the unit's performance.

SECTION 10

SHARPLES P3000-BD HORIZONTAL BOWL DECANTER CENTRIFUGE INVESTIGATION

PROCESS DESCRIPTION

Decanter centrifuges are commonly utilized in primary sludge dewatering applications in the pulp and paper industry. However, conventional decanter centrifuges have seldom been employed for biological sludge thickening due to the high sludge conditioning requirements to attain acceptable solids recoveries. The Sharples Biological Decanter (BD) series utilizes a proprietary scroll design to minimize conditioning requirements while maintaining satisfactory recoveries. The variables of importance to the performance of the Biological Decanters (BD), listed in Table 19, are the same as those that apply to conventional decanter centrifugation.

TABLE 19. SHARPLES BD DECANTER CENTRIFUGE PROCESS VARIABLES

Independent variables
1. Bowl and scroll size and configuration
2. Bowl speed
3. Pond depth
4. Scroll differential
5. Feed rate
6. Feed consistency
7. Sludge conditioning
8. Nature of the sludge solids
Dependent variables
1. Cake consistency
2. Solids recovery

DESCRIPTION OF TEST UNIT

The P3000-BD is one of the smaller production models offered by Sharples. The bowl is 14 in. (35 cm) in diameter and 30 in. (76 cm) long. The unit used in this study was operable at either 3250 or 2000 rpm resulting in centrifugal forces at the bowl wall of 2100 and 800 G's (times the force of gravity) respectively. The range of pond settings utilized was 8 to 8-3/8, and the scroll differential was varied from 2 to 20 rpm. When polymer conditioning was utilized, it was introduced in the feed zone of the bowl as is common with other types of decanter centrifuges.

OPERATION OF THE P3000-BD

In preparation for a series of test runs a bowl rpm and pond depth were selected and applied. The testing was then conducted by utilizing the desired combinations of feed rate, scroll differential, and, when applicable, polymer dosage. A stabilization period of 10 minutes was included prior to collecting samples of conditioned feed, unconditioned feed, centrate, and cake. In conjunction with sample collection the centrate flow rate was measured. Specific resistance and SVI (Sludge Volume Index) tests were conducted on conditioned and unconditioned feed samples to monitor the sludge settleability and dewaterability and to assess the effectiveness of sludge conditioning. The Sludge Volume Index test was included in the characterization of centrifuge feed because of the centrifuge's anticipated dependence upon solids settleability.

P3000-BD TEST RESULTS

Capacity

The pilot unit dewatered from 20 to 60 gpm (75 to 225 liters/min) of feed. The only available pump capable of providing such rates was a centrifugal pump, so that all data generated by the P3000-BD was on "sheared" sludge with a specific resistance ranging from about 100×10^7 to 700×10^7 sec^2/gm with typical values falling between 100×10^7 and 300×10^7 sec^2/gm. Over the same period the fresh, "unsheared" sludge specific resistance typically ranged from 50 to 150×10^7 sec^2/gm. During the course of the tests, however, the sheared and fresh sludge SVI's remained roughly equivalent, ranging between 100 and 200. Other than the large differences between conditioned and unconditioned sludges indicated in Table 20 and Figure 21, no dependence of centrifuge performance upon sludge specific resistance was observed. At least partially because the sludge volume index testing was applied to clarifier underflow rather than less concentrated mixed liquor, the sludge volume index test was

TABLE 20. CHARACTERIZATION OF CENTRIFUGE FEED

Feed	Sludge Volume Index* average/(range)	Specific resistance x 10^7 sec^2/sec average/(range)
Fresh waste activated sludge	- /(100-200)	112/57-163)
"Sheared" waste activated sludge	135/(109-217)	191/(90-722)
"Sheared" waste activated sludge conditioned with 3 to 5# polymer/ ton of solids	126/(78-169)	74/(29-132)

* Sludge consistencies of 0.6 to 0.8 percent suspended solids

not sensitive to changes in the dewaterability of the centrifuge feed. The SVI test itself becomes insensitive as consistencies approach 1 percent solids, especially on slow settling sludges. Figure 21 shows the cake consistencies associated with various feed rates at scroll differentials and pond settings yielding 90 percent solids recovery. As demonstrated by the figure, attaining 90 percent recovery at the higher feed rates resulted in lower cake consistencies. The detrimental impact of higher feed rates upon cake consistency were minimized by the application of polymer conditioning and by using higher rotational speeds. An 8-1/4 pond setting was found to be satisfactory for the range of feed rates encountered.

Cake Consistencies

The cake consistencies achieved by the P3000-BD were determined by the scroll differential and G-force. Because lower scroll differentials were required to maintain 90 percent recovery at lower feed rates, the cake consistencies attained at the lower feed rates were higher. Polymer utilization and higher G-forces probably resulted in drier cakes because these practices released additional water from the cakes and/or because lower scroll differentials were required compared to unconditioned sludges and lower G's to maintain 90 percent solids recovery

An optimum pond settling was chosen by trading off cake consistency and solids capture. A pond setting of 8-3/8 was found to be detrimental to cake consistencies while not

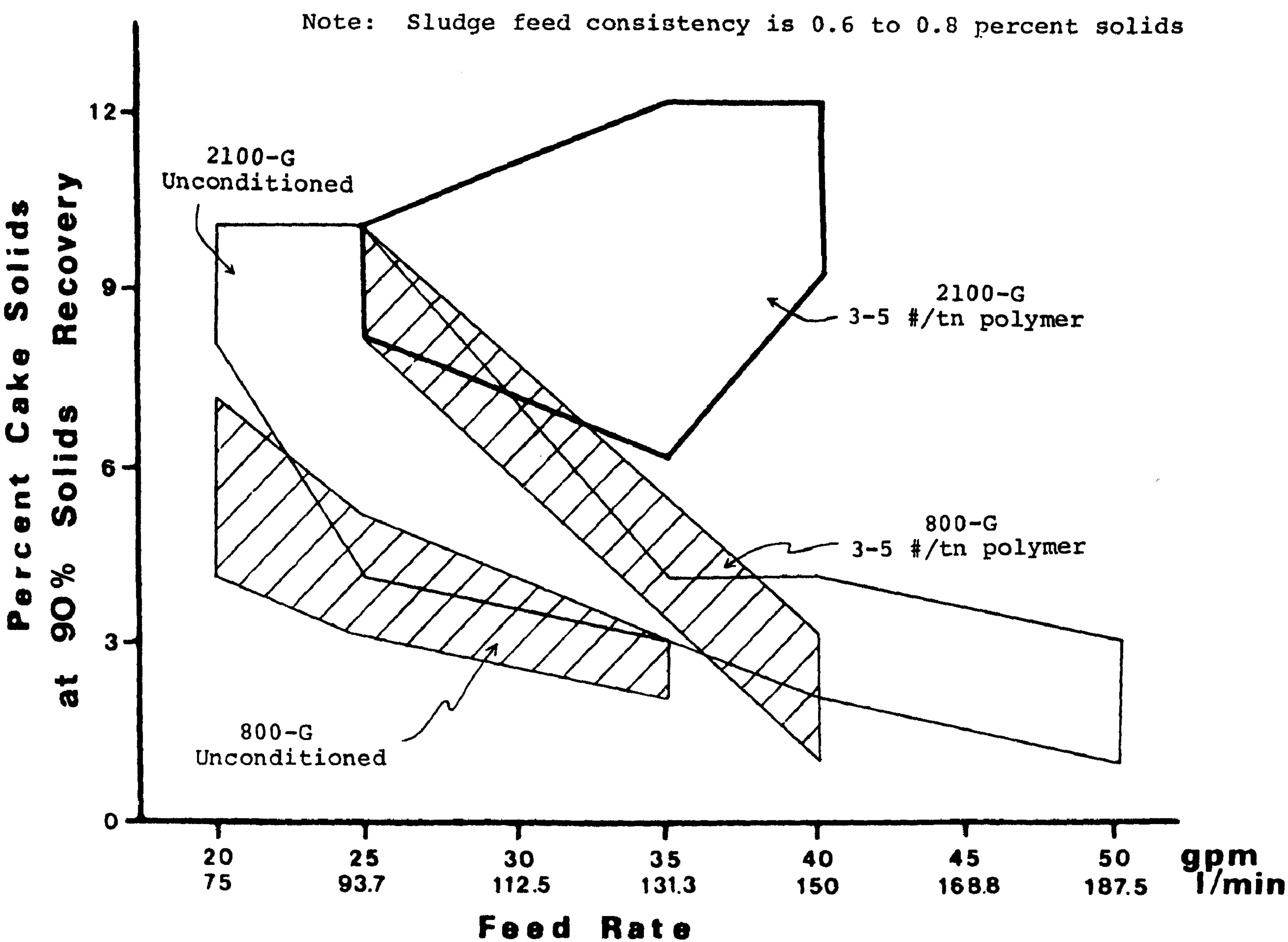


Figure 21. Centrifuge performance at 90 percent solids recovery.

resulting in substantially higher recoveries. On the other hand, a pond setting of 8 did not provide acceptable solids recovery, 8-1/4 proving to be the best compromise, as shown in Figure 22.

Solids Recoveries

Solids recoveries were maintained in excess of 90 percent by varying scroll differential. Those runs utilizing no conditioning, higher feed rates and lower centrifugal forces required higher differentials to achieve 90 percent recovery which resulted in lower cake solids. Pond settings in excess of 8-1/4 resulted in decreased cake consistencies without substantial increases in solids recoveries while those below 8-1/4 did not provide adequate recoveries.

Benefits of Sludge Conditioning

As shown in Figure 21, polymer conditioning at 3 to 5 pounds of Betz 1260 or Hercufloc 844 per ton of sludge solids (1.5 to 2.5 mg/gm) proved to be beneficial to P3000 BD performance, especially at higher feed rates and lower centrifugal forces. However, when the unit was dewatering 20 to 25 gpm (75 to 90 liters/min) at 2100-G, polymer conditioning did not result in substantially improved performance.

Overall P3000-BD Performance

At 2100-G's of centrifugal force and a pond setting of 8.25, the P3000-BD proved to be capable of thickening the "sheared", waste activated sludge from 0.8 percent consistency to 8 to 10 percent consistency. No conditioning was required to dewater 20 to 25 gpm (75 to 90 liters/min). The application of 3 to 5 pounds of Betz 1260 or Hercufloc 844 per ton of sludge solids (1.5 to 2.5 mg/gm) was advantageous in dewatering 25 to 40 gpm (90 to 150 liters/min).

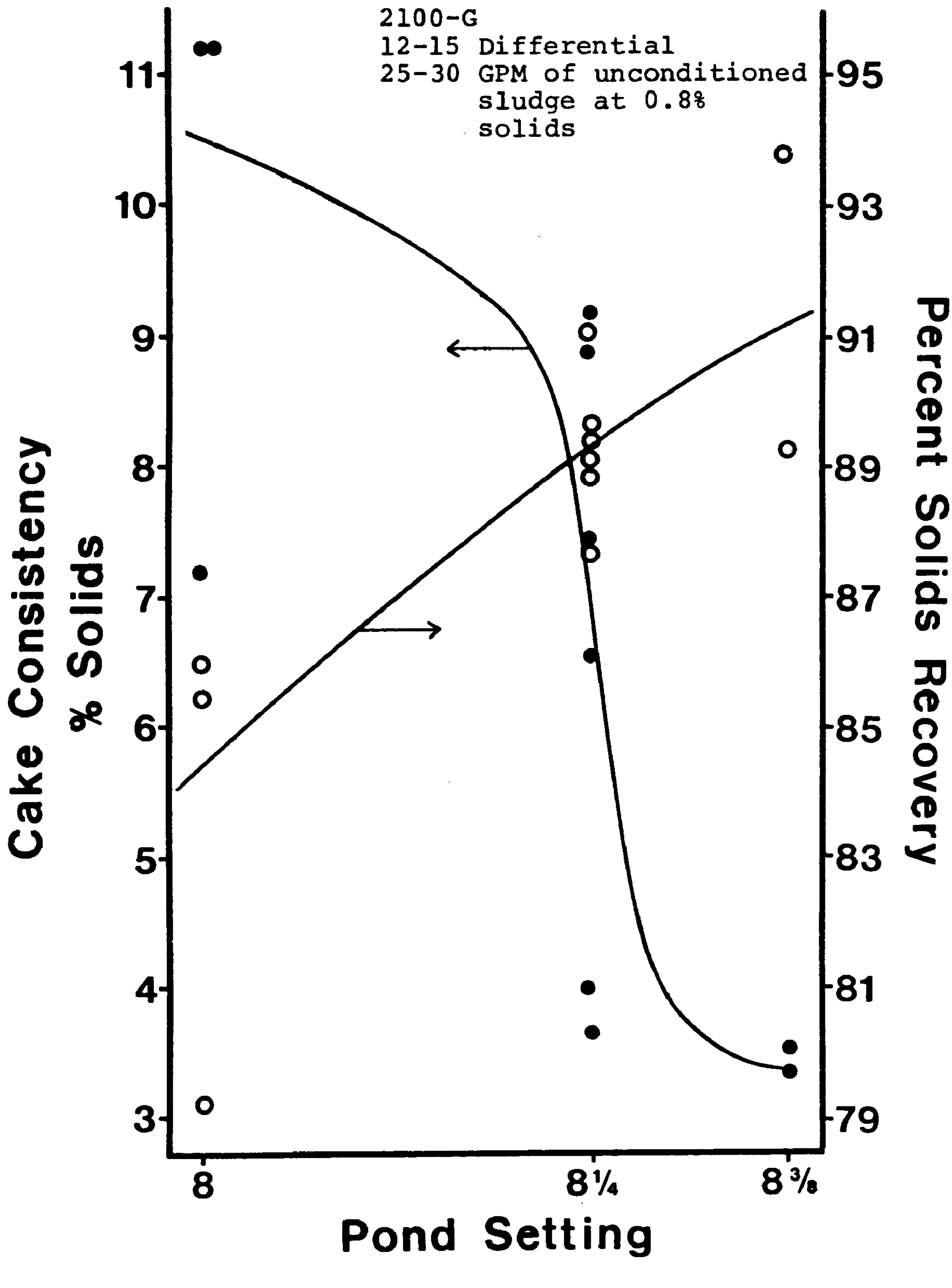


Figure 22. Effect of pond setting on solids recovery and cake consistency.

SECTION 11

ULTRAFILTER INVESTIGATION

PROCESS DESCRIPTION

Ultrafiltration is a membrane-based, separation technology. By selecting a membrane with the desired properties, and maintaining an appropriate pressure against the membrane, a separation of suspended, colloidal, or dissolved material can be accomplished. Because the separation is taking place at the membrane and there is a net movement of fluid to the membrane, there is a buildup of retained species at the membrane. The depth of this buildup, anticipated to be of major importance in sludge thickening applications, can be reduced by maintaining turbulence at the membrane surface. The variables of importance to the ultrafiltration of solids-laden feed streams are listed in Table 21.

TABLE 21. ULTRAFILTRATION PROCESS VARIABLES

Independent variables
1. Membrane characteristics
2. Membrane configuration
3. Feed rate
4. Feed consistency
5. Feed temperature
6. Operating pressure
7. Flux maintenance technique
Dependent variables
1. Flux rate
2. Solids retention
3. Pressure drop per unit membrane area
4. Concentrate consistency

DESCRIPTION OF ULTRAFILTER PILOT UNIT

The ultrafiltration unit was assembled by the National Council for Air and Stream Improvement for this study. A Westinghouse D170, 19-tube membrane module containing 9.5 ft^2 (0.91 m^2) of membrane area was selected for sludge concentration. The module contained 18 one-half inch diameter membrane tubes in series, supported upon a porous epoxy/sand material. By rotating the end caps in relation to the module, the unit could operate with as few as six tubes. Six tube operation was the mode of operation favored for this study due to pressure drop considerations.

The module was fed by a 10 gpm (37.8 liters/min) progressing cavity pump. When concentrating dilute sludge, the unit was operated on a once-through basis. To provide a constant supply of concentrated feed, the inlet of the feed pump was fitted with a 15-gallon (60 liter) feed tank into which both ultrafiltration concentrate and permeate could be recycled. Line pressure was measured at both the inlet and outlet of the module and was controlled with a throttle valve following the outlet pressure gauge. The average operating pressure was defined as the average of the inlet and outlet pressures.

OPERATION OF THE UNIT

A variety of techniques was evaluated to clean the membranes and thus maintain high flux rates. One of these was an automated backflush system designed to pump permeate back through the membrane, loosening deposited solids and restoring flux rates. The backflushing comprised from 5 to 25 percent of the total cycle time. Figure 23 is a typical flux curve generated in this study. The automated backflush system was intended to backflush the membrane as soon as the flux began to level off. By cycling the flux through only the first part of the flux curve, a higher average flux rate was anticipated. However, as Figure 24 demonstrates, by the time the backflushed volume had been recollected, the flux levels had dropped approximately to the levels attained before the backflush had been started. As a result, the backflushing technique did not result in higher overall flux rates. Since short cycle flux maintenance techniques had proven generally unsuccessful, efforts were concentrated upon finding a membrane-cleaning technique that could be applied once every 4 to 8 hours to restore flux rates to original levels. The methods tested were high velocity water feed, 5-minute backflushing, 1/2 percent Biz solution washing, sponge ball flushing of the system and combinations of these methods. Of the techniques investigated, flushing the tubular membrane system out a dozen times with a slightly oversized sponge ball was the only method capable of consistantly restoring flux levels to those attained on clean membranes.

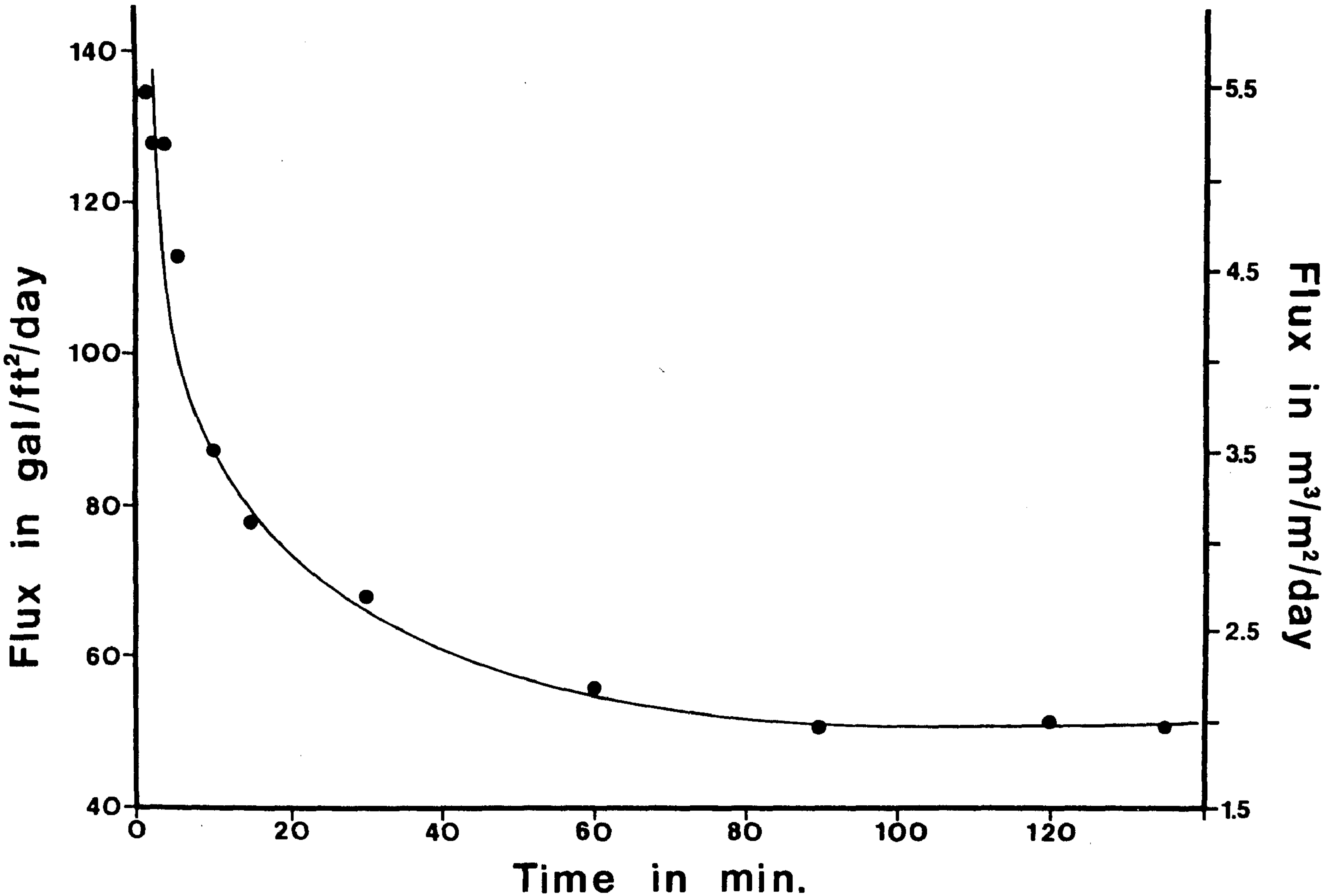


Figure 23. Typical ultrafiltration flux curve.

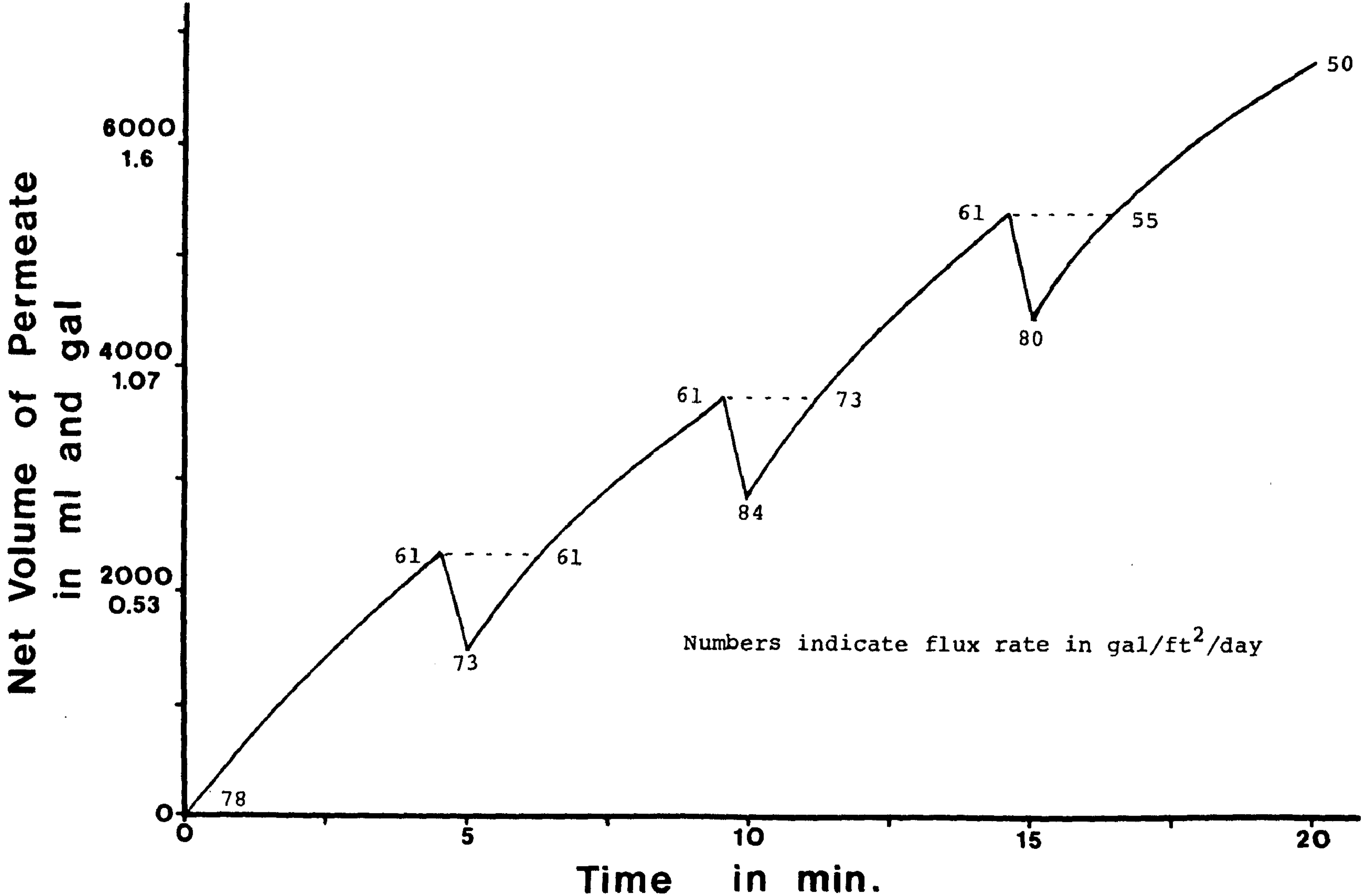


Figure 24. Effect of backflushing for flux maintenance.

After cleaning the membranes in this manner, feed was introduced at the desired rate and the average pressure in the module was adjusted. The permeate rate and temperature were measured throughout the first 5 to 10 minutes of each run. Thereafter, flux rates and permeate temperatures were measured every 30 minutes until flux remained reasonably constant for two or three consecutive readings. This constant flux rate, attained after 90 minutes in Figure 23, was defined as the steady state flux. Runs were usually terminated after the steady state flux had been determined.

Samples were taken of the feed as often as was necessary to monitor the sludge consistency, and an occasional permeate or concentrate sample was analyzed for suspended solids. When feed consistencies higher than the levels attained by gravity settling were required, the sludge concentration was increased by recirculating concentrate into the feed tank until the desired concentration was achieved and the run could be initiated. To maintain a constant feed consistency throughout these runs, both concentrate and permeate were recycled for the duration of the run, allowing continuous loading of a concentrated feed.

RESULTS OF ULTRAFILTRATION STUDY

Steady State Flux Rates

As illustrated by Figure 25, steady state flux rates were found to be strongly dependent upon the average fluid velocity in tubes (calculated by dividing the volumetric feed rate by the cross sectional area of the membrane tube) and less strongly dependent upon feed concentration.

The data in Figure 25 were generated at average operating pressures of from 25 to 75 psi ($17\text{-}52 N/cm^2$). These variations in average operating pressure did not significantly affect the flux rates attained in the range of pressures investigated.

The maximum feed rate was limited by the pressure drop through the tubes. As shown in Figure 26, the pressure drop was a function of feed rate and feed consistency, higher consistencies and feed rates resulting in larger pressure drops.

Table 22 translates the flux rates shown in Figure 25 into area requirements to accomplish various degrees of thickening at different average feed velocities. As shown, the advantage associated with thickening a feed of 2 percent consistency as opposed to 1 percent consistency is quite substantial, resulting in a reduction in required membrane area of roughly 50 percent.

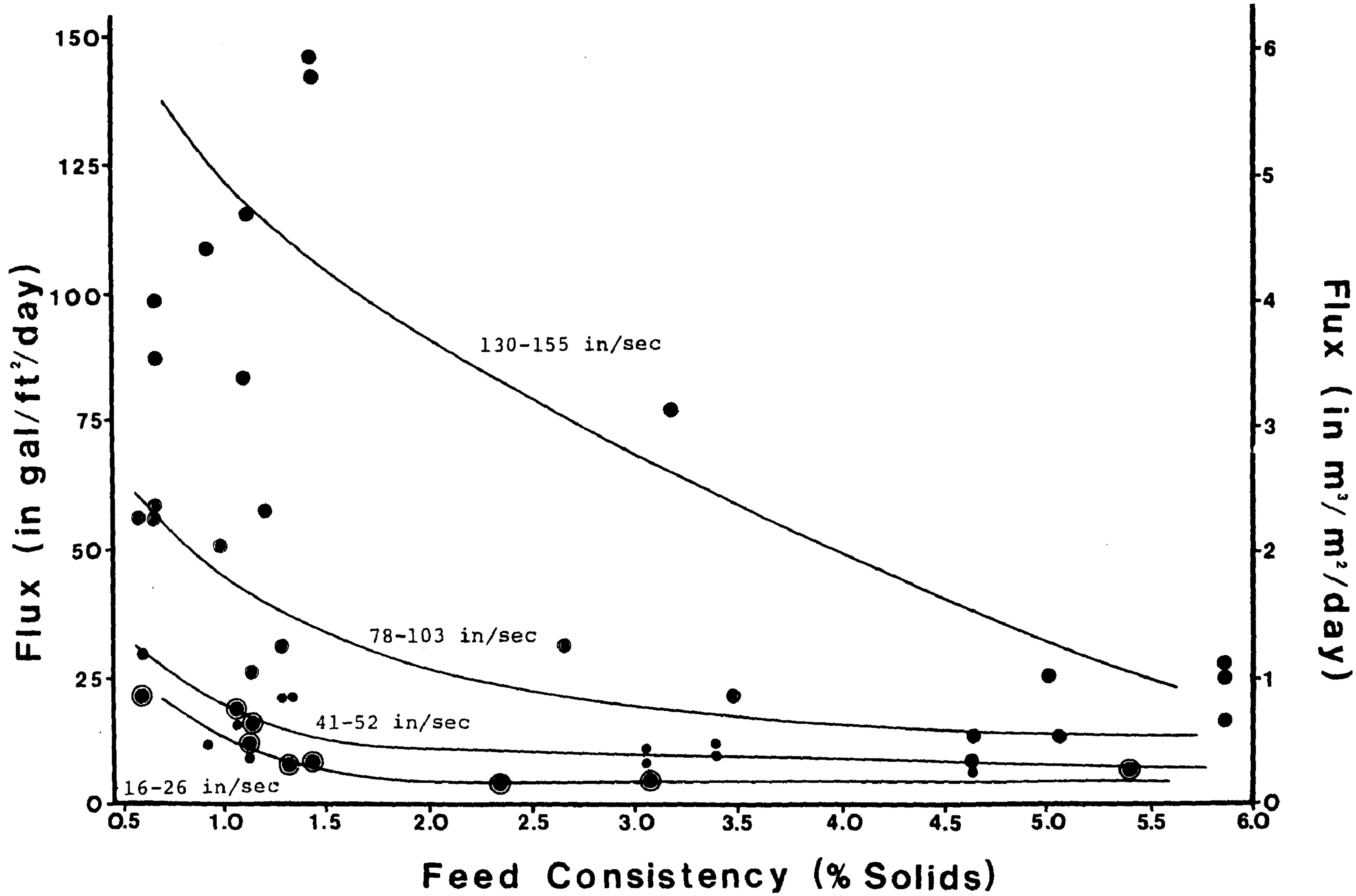


Figure 25. Ultrafiltration steady state flux rates as functions of feed consistency and fluid velocity.

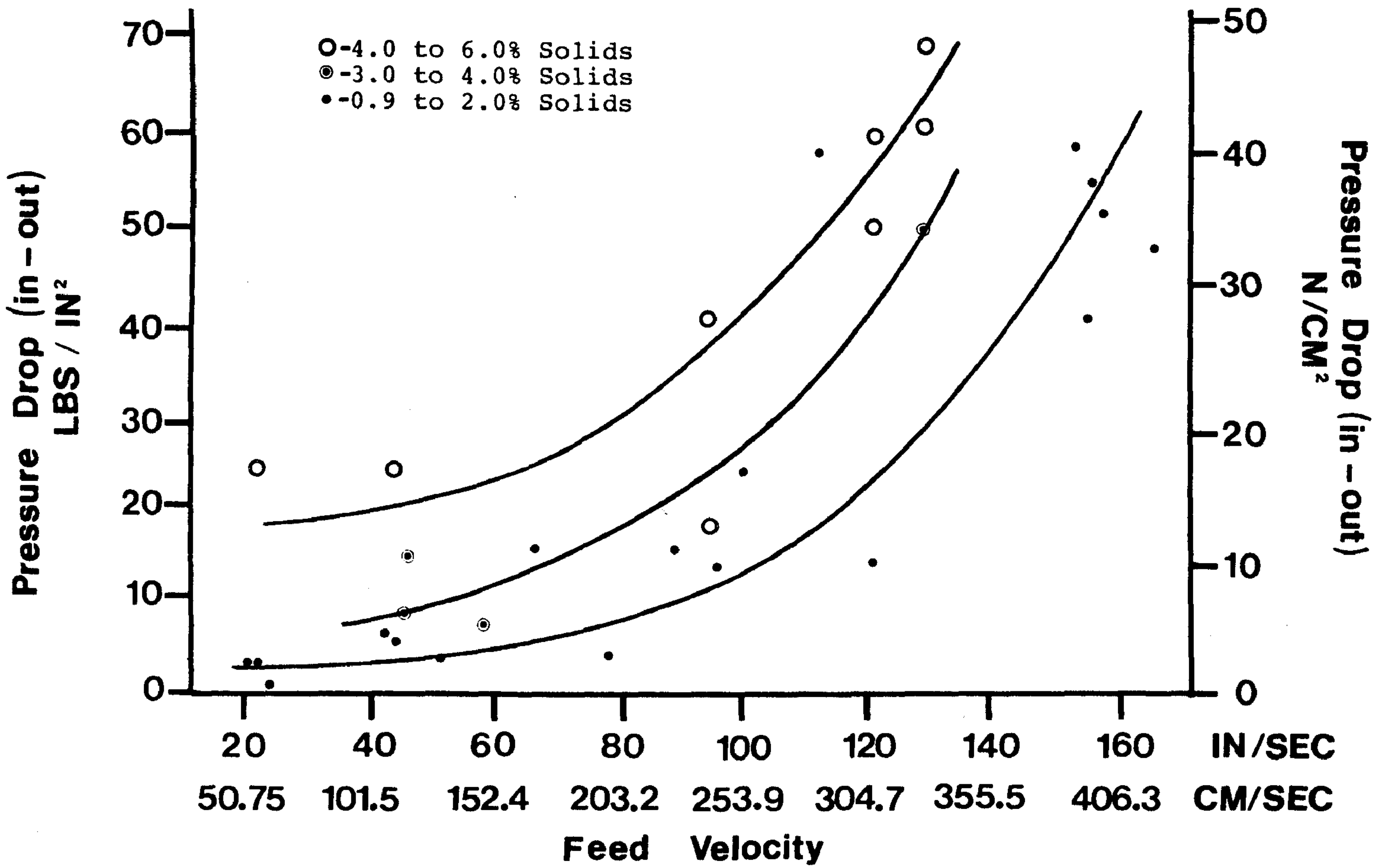


Figure 26. Pressure drop through the membrane module during six tube operation.

TABLE 22. MEMBRANE AREA REQUIREMENTS FOR 2000 POUNDS (908 KG) SLUDGE SOLIDS PER DAY

Average feed velocity of 129-155 in/sec (328-394 cm/sec)

Thickening	Flux		Area		Total area	
	$gal/ft^2/day$	$m^3/m^2/day$	ft^2	m^2	ft^2	m^2
From 1% to 2%	105	4.3	114	11.0	114	11
From 2% to 3%	80	3.3	50	4.6	164	15
From 3% to 4%	60	2.4	33	3.1	197	18
From 4% to 5%	40	1.6	30	2.8	227	21
From 5% to 6%	25	1.0	32	3.0	257	24

Average feed velocity of 78-103 in/sec (197-262 cm/sec)

Thickening	Flux		Area		Total area	
	$gal/ft^2/day$	$m^3/m^2/day$	ft^2	m^2	ft^2	m^2
From 1% to 2%	32	1.30	375	35.0	375	35
From 2% to 3%	22	0.90	182	17.0	557	52
From 3% to 4%	18	0.73	111	10.0	668	62
From 4% to 5%	15	0.61	80	7.4	748	70
From 5% to 6%	13	0.53	61	5.7	809	75

Average feed velocity of 16-51 in/sec (40-130 cm/sec)

Thickening	Flux		Area		Total area	
	$gal/ft^2/day$	$m^3/m^2/day$	ft^2	m^2	ft^2	m^2
From 1% to 2%	10	0.41	1199	111	1199	111
From 2% to 3%	8	0.33	500	47	1699	158
From 3% to 4%	8	0.33	250	23	1949	181
From 4% to 5%	7	0.29	171	16	2120	197
From 5% to 6%	7	0.29	114	11	2234	208

Table 23 demonstrates one possible module configuration to thicken sludge from 1 to 6 percent consistency. Flow velocities of about 140 in./sec (360 cm/sec) are assured by maintaining feed rates of 7 gpm (26.5 liters/min) per module. Pressure drop constraints would likely require a booster pump for every three stages over the initial stages and as much as one booster pump per stage toward the end of the thickening sequence. Lower feed velocities would allow less frequent interstage pumping but would also require much more membrane area as shown in Table 21.

Sludge feed temperatures varied from 68° to 95°F (20° to 35°C). Over that range of temperatures fresh water flux rates increased by 20 to 40 percent. Although a similar relationship was anticipated in regard to sludge flux rates, the data did not permit its identification.

Concentrate Consistencies

Using the ultrafilter, it was possible to thicken waste activated sludge to 7 percent solids by recirculating concentrate. However, as the feed consistency approached 5 percent, the pressure drop through the module became 25 to 75 psi (17 to 52 N/cm^2) depending upon the feed rate. Thickening this waste activated sludge to beyond 7 percent solids will require a membrane configuration with less pressure drop per unit membrane area.

Overall Ultrafilter Performance

The utlrafilter was capable of concentrating waste activated sludge from 1 to 7 percent consistency. The membrane area required depended upon feed velocity. The minimum estimated membrane area required per ton of solids per day to concentrate from 1 percent to 6 percent consistency was calculated to be about 260 ft^2 (24.2 m^2). One penalty encountered at high feed velocities and associated high flux rates was a very large pressure drop through each module. This suggests a potential requirement for one booster pump for each two or three module stages. Sponge flushing was found to be an effective means for restoring flux rates to levels attained with unused membranes.

TABLE 23. PROPOSED MEMBRANE CONFIGURATION TO THICKEN 1 TPD OF SLUDGE SOLIDS

Feed Velocity of 120-155 in./sec (328-394 cm/sec)
Which is Equivalent to 7 gpm per module (26.5 l/min per module)

Stage	# Modules (27 modules 254 ft^2)	Feed rate gpm	Recirculation rate, gpm	Flux	Concentrate consistency % solids
0	0	0	0	0	1.0
1	3	16.65	0	115	1.2
2	2	14.38	0	115	1.3
3	2	12.86	0	113	1.5
4	2	11.37	0	105	1.7
5	1	9.99	0	100	1.8
6	1	9.34	0	97	1.9
7	1	8.71	0	90	2.1
8	1	8.12	0	87	2.2
9	1	7.55	0	82	2.4
10	1	7.02	0	77	2.6
11	1	6.52	0	76	2.8
12	1	6.02	0	72	3.0
13	1	5.55	1.45	67	3.3
14	1	5.11	1.89	63	3.5
15	1	4.70	2.30	57	3.9
16	1	4.33	2.67	50	4.2
17	1	4.00	3.00	45	4.5
18	1	3.71	3.29	39	4.8
19	1	3.46	3.54	31	5.1
20	1	3.26	3.74	27	5.4
21	1	3.08	3.92	24	5.7
22	1	2.92	4.08	22	6.0

SECTION 12

DISCUSSION OF ALTERNATIVES

EQUIPMENT REQUIREMENTS

The equipment, conditioning, and power required to dewater 10 tpd of this mill's waste activated solids on the different dewatering technolgies investigated are shown in Table 24. Except for the centrifuge sizing calculations, which were done by Sharples Division of Pennwalt Corproation based upon this pilot study, the estimates of required capacity, sludge conditioning and cake consistency were made by Council staff based upon results of this study. The power requirements shown in the table are primarily based upon information provided by the equipment manufacturers. Those estimates made by Council staff without manufacturer consultation were those for the Squeegee, centrifuge and ultrafilter. The performance levels used to assemble Table 24 can be found in Appendix B.

FINAL SLUDGE DISPOSAL CONSIDERATIONS IN DEWATERING PROCESS SELECTION

The different dewatering technolgies investigated in this study generate cakes with varying characteristics which suggest different final disposal alternatives. The pressure filter and precoat vacuum filter are the probable alternatives if sludge incineration or long distance hauling are required. The desirability of subjecting an incinerator to the associated levels of ferric chloride, lime, or diatomaceous earth would require consideration, as would the quantities of residual ash generated in each case. The drier bulk cake consistencies and tougher cakes attained on the pressure filter might suggest its applicability in those landfill applications requiring a high degree of fill stability. However, without having determined the consolidation characteristics of the pressure filter and precoat vacuum filter cakes, an accurate assessment of their relative stability was not possible.

If final disposal requirements favor "truckable" cakes free of precoat or inorganic conditioning, the filter belt presses are probably the best suited dewatering alternatives.

TABLE 24. EQUIPMENT REQUIREMENTS FOR 10 TPD OF WASTE ACTIVATED SOLIDS BASED ON PILOT STUDY FINDINGS

Dewatering option	Equipment required	Cake solids bulk/ corrected	Conditioning required	Operating horsepower
Pressure filter, 1% feed	1 5- by 5-ft press with 80 plates	32-38/ 22-28	750# lime/ton 180# $FeCl_3$/ton	20*
Pressure filter, 2% feed	1 5- by 5-ft press with 90 plates	38-40/ 30-33	540# lime/ton 140# $FeCl_3$/ton	20*
Precoat vacuum filter, 1% feed	2 500-ft^2 filters	26-31/ 25-30	100# Diatomaceous earth/ ton	132
Precoat vacuum filter, 2% feed	1 500-ft^2 filter	26-31/ 25-30	50# Diatomaceous earth/ ton	66
Squeegee, 1% feed	75 ft of belt width	15-17	100# $FeCl_3$/ton	60
DCG-MRP, 1% feed	7 DCG-200's, 4 MRP's	15-17	5 -10# Betz 1260 or equivalent/ton	12
Tait-Andritz ADM, 1% feed	2 80-inch SDM's	17-20	8# Betz 1260 or equivalent required/ton	10
BD Centrifuge, 1% feed	2 P5400 BD's	8-10	0	120
DCG, 1% feed	7 DCG-200's	8-10	5 -10# Betz 1260 or equivalent required/ton	4
Ultrafilter, 1% feed	270 modules with 2540 ft^2 of membrane	6	0	300
Ultrafilter, 2% feed	160 modules with 1500 ft^2 of membrane	6	0	250

*Average throughout cycle

The DCG, the BD series centrifuge, and the ultrafilter appear best suited in thickening applications. Such thickening might be advantageous ahead of pressure filtration, digestion, or heat treatment or in situations where land application of semi-fluid sludge is the favored disposal alternative.

OPERATING COST CONSIDERATIONS

The various dewatering alternatives each have characteristic operating costs associated with them. Those aspects of each process which are likely to represent a major contribution to these operating costs are discussed below.

Pressure Filtration

Conditioning requirements represent the majority of the operating costs associated with this alternative. Historically, the process has also been relatively labor intensive, but many recent installations have been automated to a large degree. Power costs are less of an issue than with some of the other units evaluated in this study. The sludge conditioning chemical storage and handling facilities required for pressure filtration are probably the most elaborate of those associated with the dewatering alternatives evaluated herein. A list of pressure filter manufacturers can be found in Appendix C.

Precoat Vacuum Filtration

Precoat consumption is a crucial variable in determining the relative economies by this technology. The fact that diatomaceous earth costs can vary by plus or minus 10 percent from the west to the east coast of the United States may affect the operating costs associated with this process. Power requirements for vacuum filtration are among the highest of those estimated for the units in this study. The amount of operator attention required by this process is anticipated to be somewhat more than that associated with conventional vacuum filtration due to the necessity of precoating about once daily. The operation and maintenance of precoat storage and handling facilities also add a degree of complexity beyond conventional vacuum filtration. A list of rotary precoat filter manufacturers is included in Appendix D.

Filter Belt Presses

Sludge conditioning costs are the most significant operating costs associated with these units. Power and supervisory costs are anticipated to be among the lowest of those estimated in the study. Maintenance costs, although not well documented, are generally expected to be relatively low. The DCG-MRP system suffers a relatively poor economy of scale due to the modular nature of the equipment. A list of filter belt press manufacturers is included in Appendix E.

BD Centrifuge

The power costs anticipated for BD centrifugation, are among the highest of those estimated and are likely to be the major contributor to operating costs. Although primary sludge abrasivity has resulted in excessive centrifuge maintenance costs in some segments of the paper industry, the relatively nonabrasive nature of biological solids coupled with recent advances in hard surfacing technology indicate that the maintenance costs for BD centrifugation in this application will be substantially less than those commonly associated with primary sludge dewatering. Likely supervisory requirements are felt to be among the lowest of those encountered in this study.

Ultrafilter

The application of this alternative will require a membrane configuration which offers far less pressure drop per unit membrane area, resulting in lower power requirements for pumping. Membrane technology in general suffers a relatively poor economy of scale due to the modular nature of the membrane area offered by manufacturers.

SOLIDS RECOVERY CONSIDERATIONS

In those instances where the need for very high solids recoveries is indicated, the alternatives which suggest themselves are precoat vacuum filtration, pressure filtration, ultrafiltration, and DCG filtration.

SECTION 13

OTHER PULP AND PAPER INDUSTRY EXPERIENCE

Since the inception of this study, several pulp and paper mills, in the process of selecting sludge handling and disposal systems, have generated data on several of these dewatering technologies. Recognizing the importance of a broad data base in the selection of dewatering equipment, it is appropriate to present and discuss the data that many of these mills have made available to the NCASI.

FILTER BELT PRESSING

The pulp and paper industry has recently generated substantial amounts of pilot data and some full scale data on filter belt presses. The data that have been made available to the National Council, shown in Table 25, allow the formulation of several generalizations concerning their performance.

First, because the units are usually hydraulically limited, the most striking process compromise is between capacity, feed consistency and conditioning requirements. This suggests the desirability of prethickening sludge to minimize conditioning costs or maximize capacity. In general, these units have been applied to sludges that are very difficult to filter or centrifuge. Because of this, it is difficult to predict how they would compare with the conventional dewatering techniques on more typically fibrous primary sludge. There is some experience to suggest that filter belt presses require more sludge conditioning to provide economical throughput than is normally required in conventional vacuum filtration or decanter centrifugation where such technologies are applicable. The power requirements and cake consistencies associated with the filter belt presses, in most instances, represent significant advantages over centrifugation or vacuum filtration. The cake consistencies attained by these units on fibrous sludges are comparable to those associated with V-pressing. In general, the pilot data have shown filter belt presses to be capable of dewatering 10 to 25 gpm of primary sludge per foot of belt width, conditioned with 0 to 5 pounds of polymer per ton and generating cakes with consistencies of 30 to 40 percent solids.

These units have typically dewatered 5 to 10 gpm of

TABLE 25. PULP AND PAPER INDUSTRY FILTER BELT PRESS EXPERIENCE

Mill	Production pulping/paper	Belt press type	Sludge feed % Primary	Sludge feed % Secondary	Sludge feed % Solids	Feed rate gpm (l/m) per unit	Conditioning Type	Conditioning Amount #/tn (mg/gm)	Cake solids %	Solids recovery %	Comments
A*	Kraft/fine	20-inch Tait- Andritz, S-press	100	0	5.0	35 (130)	Betz 1260	3.0 (1.5)	35	95	
A	Kraft/fine	80-inch Tait- Andritz, P-press	100	0	5.0	120 (450)	Betz 1260	2.0 (1.0)	38	95	
B	Sulfite/	20-inch Tait- Andritz, S-press	100	0	2.0	70 (260)	-	0	22	99	Ray cells + fiber fines
B	Sulfite/	20-inch Tait- Andritz, P-press	100	0	2.0	70 (260)	-	0	36	99	Ray cells + fiber fines
C	/Tissue	Two 76"x101" Smith & Loveless	100	0	3.5	12 (45)	Polyelectro- lyte	30.0 (15.0)	10	99	Fiber fines
D	Kraft/linerboard	Two 2-meter Dravo DJ sinus	0	100	4.5	25 (95)	Polyelectro- lyte	30.0 (15.0)	12	98	Appreciable quantity of fiber in sludge
E	/paperboard	Two 1-meter Dravo DJ sinus	0	100	7.0	17 (65)	Polyelectro- lyte	20.0 (10.0)	19	99	
F*	NSSC/NSSC board	20-incn Tait- Andritz, S-press	0	100	2.0	15 (55)	Polyelectro- lyte	20.0 (10.0)	12	98	
G*	Kraft/kraft	20-inch Tait- Andritz, P-press	100	0	7.0	50 (190)	-	0	30	98	
G*	/Coated groundwood	20-inch Tait- Andritz, P-press	100	0	4.0	15 (55)	Polyelectro- lyte	4.0 (2.0)	33	99	
H*	Kraft/linerboard	20-inch Tait- Andritz, P-press	100	0	3.0	16 (60)	Polyelectro- lyte + alum	2+250 (1+125)	25	98	
H*	Kraft/linerboard	20-inch Tait- Andritz, P-press	75	25	4.5	16 (60)	Polyelectro- lyte	10.0 (5.0)	24	98	
H*	Kraft/linerboard	20-inch Tait- Andritz, S-press	0	100	2.0	16 (60)	Polyelectro- lyte + alum	10+100 (5+50)	17	95	
H*	Kraft/linerborad	24-inch Passe- vant vacupress	100	0	3.0	20 (75)	Polyelectro- lyte	2.5 (1.25)	19	95	

* Pilot data

(continued)

TABLE 25 (continued)

Mill	Production pulping/paper	Belt press type	Sludge feed			Feed rate gpm (l/m) per unit	Conditioning		Cake solids %	Solids recovery %	Comments
			% Primary	% Secondary	% Solids		Type	Amount #/tn (mg/gm)			
I*	/fine	20-inch Tait-Andritz, S-press	100	0	2.0	30 (115)	-	0	25	91	
J*	Deinking/fine	20-inch Tait-Andritz, S-press	90	10	14.0	5 (20)	Polyelectro-lyte	1.5 (0.75)	34	99	
J*	Deinking/fine	20-inch Tait-Andritz, S-press	0	100	2.5	1 (4)	Polyelectro-lyte	6.0 (3.0)	10	99	
K*	Kraft/fine	20-inch Tait-Andritz, S-press	100	0	4.0	9 (35)	-	0	30	93	
L*	Kraft/fine	20-inch Tait-Andritz, S-press	100	0	2.5	20 (75)	Polyelectro-lyte	1.0 (0.5)	30	99	
L*	Kraft/fine	20-inch Tait-Andritz, S-press	85	15	2.0	10 (40)	Polyelectro-lyte	2.5 (1.25)	22	99	
M*	Kraft/fine	20-inch Tait-Andritz, P-press	100	0	3.0	30 (115)	Betz 1260	5.0 (2.5)	40	98	
N*	Kraft/fine	20-inch Tait-Andritz, P-press	0	100	4.0	5 (20)	Betz 1260	18.0 (9.0)	14	98	Anaerobic alum coagulated ASB solids
O*	Sulfite/specialties	20-inch Tait-Andritz, P-press	100	0	3.0	40 (150)	Polyelectro-lyte	4.0 (2.0)	35	99	
P*	/fine	20-inch Tait-Andritz, P-press	80	20	3.0	30 (115)	Hercufloc 859	1.0 (0.5)	35	98	
P*	Groundwood/specialties	20-inch Tait-Andritz, S-press	80	20	3.0	30 (115)	Hercufloc 859	1.0 (0.5)	20	98	
Q*	Groundwood/specialties	20-inch Tait-Andritz, P-press	see comment		1.5	20 (75)	Polyelectro-lyte	5.0 (2.5)	35	97	Lagoon bottom solids
Q*	Groundwood/specialties	20-inch Tait-Andritz, P-press	0	0	5.0	7 (25)	Polyelectro-lyte	7.0 (3.5)	30	80	Alum filter plant sludge
R*	Kraft/linerboard	20-inch Tait-Andritz, S-press	100	0	2.0	15 (55)	Betz 1260	10.0 (5.0)	25	97	Occasionally anaerobic
R*	Kraft/linerboard	0.5-meter Carter press	100	0	2.0	40 (150)	Polyelectro-lyte	10.0 (5.0)	30	99	Occasionally anaerobic

* Pilot data

biological sludge per foot of belt width, conditioned with 10 to 20 pounds of polymer per ton, and attained cake consistencies of 12 to 19 percent solids. Performance levels on combined sludges have fallen between those associated with primary sludges and biological sludges.

Because of the limited amount of full scale operating experience with filter belt presses in this country, the operating problems encountered to date are largely associated with startup and refinement of machinery. The more common difficulties include bearing failures, belt tracking problems and occasional belt tearing. Depending upon the unit configuration, utilization of high belt tensions or nip pressures may suggest inclusion of a screening device to remove large pieces of debris from the feed. The limited amount of comparative pilot and full scale data show generally good agreement between manufacturer's performance estimates based upon pilot work at mill sites and full scale performance.

As of July 1976, at least 12 mills in the United States had purchased filter belt presses.

Appendix E includes those manufacturers of moving belt presses identified by the National Council.

PRESSURE FILTRATION

Of the emerging technologies evaluated in this study, pressure filtration was initially the most commonly applied alternative. As a result, full scale as well as pilot data are available. The pilot and full scale pressure filter data provided to the National Council are shown in Table 26. In general, the great variation in reported loading rates can be attributed to variations in feed consistencies, the amount of admix utilized, and the nature of the sludge solids. The mills having evaluated both 100 and 200 psi units have generally selected lower pressure units, citing a loss of less than 5 percent cake consistency in exchange for greatly reduced capital costs. However, the exceptions to this generalization provide evidence that the optimum operating pressure may in fact vary from sludge to sludge. A universal necessity for precoat utilization has not been supported by industry experience. Several mills having operated pressure filters continuously for anywhere from several months to three years report acceptable performance without precoat utilization. These mills apply thorough, periodic media cleaning (weekly to monthly). However, as was the case with optimum operating pressure, it is likely that the relative advantages of precoat utilization will depend upon the nature of the solids being dewatered and the type of filter press involved.

Industry experience has shown that startup problems with

TABLE 26. PULP AND PAPER INDUSTRY PRESSURE FILTER EXPERIENCE

Mill	Production pulp/paper	Manufact-urer	Size of unit(s)	% Sec-ondary solids	Feed consis-tency % solids	Conditioning requirements	Total solids dewatered lb/hr/ft^2 (kg/hr/m^2)	% Cake solids	Comments
A	Kraft/fine	Passevant	100, 52-inch chambers	0	4.0	7% lime	2.8 (14.0)	35	
A*	Kraft/fine	Passevant	100, 52-inch chambers	20	-	15% lime + diato-maceous earth pre-coat	3.6 (18.0)	35	
B	Kraft/kraft	Passevant	75, 62-inch chambers	10	4.0	3.5# polymer/ton + coke breeze pre-coat	0.6 (3.0)	35	Includes 25% alum color removal solids and 25% silica
C	Kraft/fine	Netzsch	315, 47x47-inch chambers	0	2.5	1.5 # polymer/ton	1.3 (6.0)	50	
D	Kraft/linerboard	Passevant	64, 62-inch chambers	100	11.0	2# flyash/# sludge + flyash precoat	3.1 (15.0)	50	
E	Thermomechanical/	Edwards-Jones	68, 48x48-inch chambers	50	2.5	10% alum	0.3 (1.5)	30	
F	Kraft/fine	Shriver-Johnson	146, 48x72-inch chambers	20	4.0	7% lime+1-1/2% $FeCl_3$	1.2 (6.0)	37	
G*	Groundwood, kraft/ newsprint	Edwards-Jones	109, 48x72-inch chambers	15	2.0	5% alum	1.4 (7.0)	38	
H*	/paperboard	Passevant	120, 62-inch chambers	100	2.0	0.5 lb lime/lb sludge + 6% $FeCl_3$ + cement dust precoat	0.4 (2.0)	38	
I*	Kraft/fine	Edwards-Jones	250, 60-inch plates	20	8.0	lime + $FeCl_3$	2.5 (12.0)	43	

*Pilot data

pressure filters can be substantial. Among the more common difficulties encountered are (a) plate warping or breakage (most commonly associated with plastic plates), (b) media tearing, (c) media blinding, (d) poor sludge and/or precoat distribution within the chambers, and (e) malfunctions with timing and plate shifting devices. In most instances these problems have been resolved, some, however, only after months (in one instance over a year) of more or less trial and error trouble shooting. It is yet too early to accurately estimate the amount of operator attention that these units will require once startup difficulties have been dealt with successfully. Although the data is sparce, the industry experience presently available suggests that pilot data can, with experienced interpretation, provide a reasonably accurate estimate of full scale capacity requirements. However, the sensitivities (reported by several mills and confirmed in this study) of pressure filtration to day-to-day variations in feed consistency and the nature of the sludge solids suggest the application of conservative estimates in filter press sizing. This is especially true since presses purchased with room for additional plates can accomodate an increase in capacity at a relatively low cost compared to the purchase of additional presses.

As of July 1976, at least 11 mills in the United States had purchased pressure filters for sludge dewatering.

The manufacturers of the pressure filters that have been applied to pulp and paper industry sludges are listed in Appendix C.

REFERENCES

1. Miner, R.A., D.W. Marshall, and I. Gellman. Sludge Dewatering Practice in the Pulp and Paper Industry. NCASI Technical Bulletin No. 286. NCASI, New York, New York, June 1976.

2. Eckenfelder, W. Wesley. Industrial Water Pollution Control. McGraw-Hill Book Company, 1966. p. 236.

3. Process Design Manual for Sludge Treatment and Disposal. U.S. Environmental Protection Agency Technology Transfer Series, EPA 625/1-74-006, U.S. Environmental Protection Agency, 1974. pp. 7-2 through 7-11.

4. Martin, James., and Phillip L. Hayden. Pressure Filtration of Waste Industrial Sludges, Modeling, Design, and Operational Parameters. Presented at the 31st Annual Purdue Industrial Waste Conference, May 1976.

5. Adams, Carl E., and W. Wesley Eckenfelder, ed. Process Design Techniques for Industrial Waste Treatment. Enviro Press, Nashville, Tennessee and Austin, Texas, 1974. p. 170.

6. Smith, Gordon R.S. How to Use Rotary, Vacuum, Precoat Filters. Chemical Engineering, 83 (4):84, February 16, 1976.

7. Carpenter, W.L., and I. Gellman. Evaluation of Fly Ash as a Filter Aid for Precoat Vacuum Filtration of Papermill Sludges. NCASI Technical Bulletin No. 158. NCASI, New York, New York, July 1962.

8. Lippert, R.E., and M.C. Skriba. Evaluation and Demonstration of the Capillary Suction Dewatering Device. Environmental Protection Agency Technology Series, EPA-670/2-74-017. Environmental Protection Agency, Cincinnati, Ohio, 1974.

APPENDIX A

PILOT EQUIPMENT PERFORMANCE DATA

TABLE A-1. PRECOAT VACUUM FILTER DATA

Drum speed RPM	Drum submergence %	Vacuum " Hg	Knife advance mil/min	Feed consistency % solids	Bulk cake consistency % solids	Solids loading rate $lb/ft^2/hr$	Sludge specific resistance $\times 10^7$ sec^2/gm	Diatomaceous earth content in cake - % of solids in cake
1.0	25	23	2.00	0.96	32.7	0.42	168	32
0.5	25	23	1.00	0.96	32.3	0.41	168	20
1.0	25	23	2.00	0.62	31.2	0.75	104	21
1.0	40	23	2.00	0.62	26.9	0.89	104	18
2.0	40	23	4.00	0.62	24.6	0.72	104	36
0.5	40	23	0.75	1.40	29.1	1.19	104	6
2.0	25	23	3.00	1.40		0.17	104	64
0.5	25	23	1.00	1.40	33.6	0.71	104	12
1.0	25	23	2.00	0.78	27.7	0.69	85	22
1.5	25	23	3.00	1.40	28.9	1.69	85	15
0.5	25	23	1.00	1.40	30.0	0.73	85	12
0.5	40	23	0.75	1.40	25.8	1.46	85	5
1.0	40	23	2.00	1.40	23.4	2.09	85	9
1.0	25	23	3.00	1.30	31.8	0.69	490	30

(continued)

TABLE A-1 (continued)

Drum speed RPM	Drum submer-gence %	Vacuum " Hg	Knife advance mil/min	Feed consis-tency % solids	Bulk cake consistency % solids	Solids loading rate $lb/ft^2/hr$	Sludge specific resistance $\times 10^7$ sec^2/gm	Diatomaceous earth content in cake - % of solids in cake
0.5	25	23	1.50	1.50	32.3	0.42	490	26
0.5	40	23	2.00	1.60	20.1	0.60	790	25
0.2	40	23	0.75	1.60	24.3	0.36	790	17
0.5	33	23	1.00	1.60	29.3	0.55	790	15
1.0	25	23	2.00	0.61	32.0	0.66	84	23
1.5	25	23	5.00	0.61	32.6	0.81	85	38
0.5	25	23	1.00	0.57	32.3	0.38	86	21
0.5	33	23	1.00	0.57	32.1	0.33	86	23
1.0	33	23	2.50	0.57	37.4	0.60	79	29
1.0	25	23	3.00	0.30	32.3	0.48	72	38
0.5	25	23	0.50	0.42	31.9	0.34	72	13
1.0	40	23	2.00	0.71	26.7	1.40	72	12
0.5	40	23	0.50	0.81	27.2	0.96	40	5
0.5	33	23	0.50	0.81	29.8	0.82	40	6

(continued)

TABLE A-1 (continued)

Drum speed RPM	Drum submergence %	Vacuum " Hg	Knife advance mil/min	Feed consistency % solids	Bulk cake consistency % solids	Solids loading rate $lb/ft^2/hr$	Sludge specific resistance $\times 10^7$ sec^2/gm	Diatomaceous earth content in cake - % of solids in cake
1.00	33	23	1.0	2.27	23.3	2.82	181.0	3
1.00	25	23	1.0	1.51	32.6	1.35	71.5	7
0.50	25	23	0.5	1.18	33.2	0.55	76.5	8
0.50	40	23	0.5	0.87	29.7	0.62	85.4	7
1.00	40	23	1.0	1.11	22.8	1.27	82.5	7
0.50	33	22	0.5	1.80	29.3	1.25	94.3	4
0.50	33	15	0.5	2.10	27.3	1.04	78.6	5
0.50	33	10	0.5	1.80	26.1	1.02	106.7	5
0.50	33	7.5	0.5	2.00	24.6	1.42	157.0	3
0.67	25	22	2.0	1.53	36.2	0.57	185.0	26
0.50	25	22	0.5	1.60	33.1		205.0	
0.50	40	22	1.0	1.32	28.0	0.55	224.0	15
0.67	40	23	1.0	1.43	26.8	0.65	238.0	13
0.50	25	18	0.5	1.66	33.1	0.89	179.0	5
1.00	25	18	2.0	1.14	33.4	1.06	90.0	16
1.00	40	18	2.0	0.61	34.1	0.86	94.0	19
0.50	40	18	0.5	0.75	29.4	0.61	105.0	8

(continued)

TABLE A-1 (continued)

Drum speed RPM	Drum submergence %	Vacuum " Hg	Knife advance mil/min	Feed consistency % solids	Bulk cake consistency % solids	Solids loading rate $lb/ft^2/hr$	Sludge specific resistance $\times 10^7$ sec^2/gm	Diatomaceous earth content in cake - % of solids in cake
1.00	33	23.0	0.75	0.81		1.22	40	6
1.00	25	24.0	1.50	0.85	32.0	0.85	140	15
0.50	25	24.0	0.75	0.95		0.56	111	12
0.50	40	24.0	1.00	1.14	26.2	1.12	70	8
1.00	40	24.0	1.00	1.11	22.3	1.35	82	7
1.00	40	24.0	1.50	1.00	25.8	1.05	126	12
1.00	25	7.50	1.50	1.50	24.2	1.36	90	10
1.00	25	10.0	1.50	1.64	25.6	1.78	90	8
1.00	25	15.0	1.50	1.60	26.5	1.68	139	8
1.00	25	24.0	1.50	1.46	28.8	1.57	120	9
1.00	25	23.0	1.00	1.89	28.0	2.12	92	4
1.00	25	23.0	0.50	2.20	27.2	1.45	146	3
0.13	40	23.0	0.50	2.53	28.0	0.76	201	6
1.00	40	23.0	1.00	1.47	21.1	2.00	132	5
0.50	40	23.0	0.50	1.99	23.0	1.78	137	3

TABLE A-2 (Part I). DUAL CELL GRAVITY FILTER (DCG) - MULTIPLE ROLL PRESS (MRP) DATA
(Fresh Waste Activated Sludge)

DCG belt speed in/min	DCG belt mesh	Feed cons. % solids	Conditioning type	Conditioning amount - % of sludge solids	Conditioned feed specific resistance $\times 10^7$ sec^2/gm	Unconditioned feed specific resistance $\times 10^7$ sec^2/gm	Feed rate gpm	DCG cake cons. % solids	DCG solids recovery - %	MRP belt speed in/min	MRP roll pressures lb/linear in					MRP cake cons. % solids	MRP solids recovery - %
											First	Second	Third	Fourth	Final		
135	100	0.9	Betz 1260	65.0			15	8.7	99.0		7	7	7	7	35	13.6	97.8
135	100	0.9	Betz 1260	10.6	2.5	48.3	10	8.6	98.3	168	7	7	7	7	35	13.5	98.8
57	100	0.9	Betz 1260	10.6	1.5		10	8.6	99.5	168	7	7	7	7	35	13.6	99.3
135	100	0.9	Betz 1260	5.6	80.0		10	7.5	99.5	168	7	7	7	7	35	12.7	86.9
135	100	1.0	Betz 1260	20.0	6.8		10	8.3	98.5	168	7	7	7	7	35	13.2	92.6
57	100	1.0	Betz 1260	20.0	6.5		10	8.7	99.5	168	7	7	7	7	35	12.7	92.6
135	100	1.0	Betz 1260	11.0	7.2		15	7.5	100.0	168	7	7	7	7	35	13.6	92.8
135	100	0.7	Betz 1260	13.5	10.2		5	8.5	97.2	168	7	7	7	7	35	13.5	81.3
57	100	0.7	Betz 1260	13.5	13.5		5	8.0	99.0	168	7	7	7	7	35	13.4	80.8
135	100	0.8	Betz 1260	6.1	27.7		5	7.2	99.0	168	7	7	7	7	35	13.6	79.5
57	100	0.9	Betz 1260	5.4	40.1		5	7.0	99.0	168	7	7	7	7	35	13.4	95.4
135	100	1.0	Betz 1260	21.0	2.5	239.9	5	6.8	99.0	168	7	7	7	7	35	12.8	96.2
57	100	1.0	Betz 1260	20.0	7.2		5	8.3	97.8	168	7	7	7	7	35	12.7	89.1
135	100	1.0	$FeCl_3$	5.0	13.1		5	6.1	99.0	168	7	7	7	7	35	15.1	92.2
135	100	1.1	$FeCl_3$	4.5	11.4		10	5.3	99.0	168	7	7	7	7	35	15.5	
135	40	0.8	Betz 1260	13.0			10	9.5		168	7	7	7	7	52	16.5	
135	40	0.8	Betz 1260	13.0			10	9.7		84	7	7	7	7	52	17.0	
135	40	0.8	Betz 1260	13.0			10	9.3		396	7	7	7	7	52	14.9	
135	40	0.8	Betz 1260	13.0	7.0		10	9.8		396	7	14	21	29	52	15.8	
135	40	0.8	Betz 1260	13.0			10	9.4		168	7	14	21	29	52	18.3	
135	40	0.8	Betz 1260	13.0			10			168	7	14	21	29	87	17.9	
135	40	0.8	Betz 1260	13.0			10	9.0		396	7	14	21	29	87	15.7	
135	40	0.8	Betz 1260	13.0			10	9.2		168	29	29	35	35	87	17.5	
135	40	0.8	Betz 1260	33.0			10	8.6		168	7	14	21	29	52	16.8	
135	40	0.8	Betz 1260	12.5	2.0	35.8	10	9.5		168	7	14	21	29	52	15.7	98.4
135	40	0.8	Betz 1260	12.5			10	8.8		168	7	14	21	29	87	16.0	96.2
135	40	0.8	Betz 1260	12.5			10	9.0		396	7	14	21	29	87	15.0	96.2
135	40	0.8	Betz 1260	12.5			10			396	29	29	35	35	87	15.5	95.1
135	40	0.8	Betz 1260	3.1	15.6		10	7.7		168	29	29	35	35	87	14.3	88.8
135	40	0.8	Betz 1260	3.1			10	7.2		168	7	14	21	29	87	14.1	89.5

TABLE A-2 (Part II). DUAL CELL GRAVITY FILTER (DCG) - MULTIPLE ROLL PRESS (MRP) DATA
(Degraded Waste Activated Sludge)

DCG belt speed in/min	DCG belt mesh	Feed cons. % solids	Conditioning type	Conditioning amount - % of sludge solids	Conditioned feed specific resistance x10^7 sec^2/gm	Unconditioned feed specific resistance X10^7 sec^2/gm	Feed rate gpm	DCG cake cons. % solids	DCG solids recovery - %	MRP belt speed in/min	MRP roll pressures lb/linear in: First	Second	Third	Fourth	Final	MRP cake cons. % solids	MRP solids recovery - %
135	100	1.0	Betz 1260	20.0	1.2	233.8	5	8.1	99.0	168	7	7	7	7	35	12.9	98.9
135	100	0.9	Betz 1260	11.0	16.0		5	8.2	99.0	168	7	7	7	7	35	12.3	98.2
135	100	0.9	Betz 1260	5.5	38.2		5	8.3	98.8	168	7	7	7	7	35		97.0
135	100	0.8	Betz 1260	12.5	16.5	355.9	10	8.0	99.0	168	7	7	7	7	35	13.8	96.0
135	100	0.8	Betz 1260	6.2	→	→	10	overflow		→	→	→	→	→	→	→	→
135	100	1.0	Betz 1260	20.0	1.8		10	7.8	98.9	168	7	7	7	7	35	13.4	97.6
135	100	1.0	Betz 1260	10.0	→	→	15	overflow		→	→	→	→	→	→	→	→
135	100	0.9	Betz 1260	22.0	3.4		15	6.8	99.0	168	7	7	7	7	35	12.8	98.1
135	100	0.7	Betz 1260	28.6	5.5	300.0	15	8.0	99.0	168	7	7	7	7	35	13.2	97.7
135	100	0.9	Hercufloc 859	22.0	1.8		15	8.6	99.0	168	7	7	7	7	35	14.4	99.0
135	100	0.8	Hercufloc 859	18.7	3.5		15	8.1	99.0	168	7	7	7	7	35	15.0	97.8
135	40	0.8	Betz 1260	26.4	7.4	181.3	15	9.2	98.9	168	7	7	7	7	52	15.0	81.0
135	40	1.0	Betz 1260	15.8	10.4	112.7	15	8.5	99.0	168	7	7	7	7	52	15.6	92.8
135	40	1.0	Betz 1260	18.0	→	→	22	overflow		→	→	→	→	→	→	→	→
135	40	1.0	Betz 1260	19.8	→	→	20	overflow		→	→	→	→	→	→	→	→
135	40	1.0	Betz 1260	22.0	4.8	139.0	18	7.4	99.0	168	7	7	7	7	52	16.8	93.5
135	40	0.8	Betz 1260	19.8	8.7	174.4	10	9.6	99.0	168	7	7	7	7	52	17.3	90.1
57	40	0.8	Betz 1260	19.8	8.4		10	9.5	98.0	168	7	7	7	7	52	16.9	84.0
135	40	0.7	Betz 1260	3.8	19.4		10	8.9	97.5	168	7	7	7	7	52	15.8	83.0
135	40	0.9	Betz 1260	11.7	6.6		15	9.8	98.9	168	7	7	7	7	52	16.0	95.3
135	40	0.7	Betz 1260	3.8	24.6		15	9.4	99.0	168	7	7	7	7	52	16.5	91.8
135	40	0.8	Betz 1260	24.7	7.7		11	7.7	99.0	168	7	7	7	7	52	17.0	95.3
135	40	1.2	Betz 1260	16.7	1.1	162.1	10	8.9	99.0	168	7	14	21	29	52	15.8	95.7
135	40	1.2	Betz 1260	4.2	→	→	10	overflow		→	→	→	→	→	→	→	→
135	40	1.2	Betz 1260	8.4	37.3		10			168	7	14	21	29	52	15.4	93.4
57	40	1.2	Betz 1260	8.4	→	→	10	overflow		→	→	→	→	→	→	→	→
70	40	1.2	Betz 1260	8.4			10	8.5	99.0	168	7	14	21	29	52	16.2	93.9
135	40	1.2		0			10		75.0 ← before overflow								

TABLE A-3. SHARPLES BD-3000 CENTRIFUGE DATA

Centrifugal force x gravity	Pond setting	Scroll differential RPM	Feed rate GPM	Feed consistency % solids	Solids recovery %	Cake consistency % solids	Unconditioned sludge volume index	Conditioned sludge volume index	Unconditioned sludge specific resistence $x10^7$ sec^2/gm	Conditioned sludge specific resistence $x10^7$ sec^2/gm	Conditioning type	Conditioning amount lb/tn
2100	8.00	20	26.5	0.66	82.6	9.8	130		230			0
2100	8.00	15	26.3	0.64	86.0	11.2	143		525			0
2100	8.00	20	26.2	0.69	72.8	11.2						0
2100	8.00	15	26.4	0.75	79.2	11.2	128		395			0
2100	8.00	10	26.1	0.66	83.7	11.8	149					0
2100	8.00	6	26.1	0.53	86.7	11.2						0
2100	8.00	2	25.9	0.70	69.2	13.5	135		180			0
2100	8.25	20	28.4	0.61	93.1	4.7	138		191			0
2100	8.25	15	23.9	0.83	89.6	9.2	115		201			0
2100	8.25	10	27.2	0.69	89.0	7.7	123		244			0
2100	8.25	6	26.4	0.77	80.9	11.5	114		179			0
2100	8.25	2	25.9	0.68	69.6	13.8	125		202			0
2100	8.00	6	41.2	0.66		8.6			149			0
2100	8.00	8	39.1	0.62	80.8	6.3			126			0
2100	8.00	10	42.1	0.64	90.6	4.0			120			0
2100	8.00	10	26.8	0.64	81.6	7.5			181			0
2100	8.00	12	27.2	0.68	85.4	7.2			139			0
2100	8.00	14	27.0	0.48		6.8			139			0
2100	8.25	8	19.2	0.74	90.5	10.9	126		201			0
2100	8.25	8	26.6	0.66	91.2	9.8	141		181			0
2100	8.25	15	19.6	0.77	92.6	8.8	125		162			0
2100	8.25	8	27.0	0.83	87.1	9.8	114		112			0
2100	8.25	20	19.4	0.77	92.9	9.9						0
2100	8.25	15	19.1	0.71	89.9	10.6	136		155			0
2100	8.25	20	29.4	0.75	92.5	4.6						0
2100	8.25	8	19.3	0.80	90.8	11.0	120		261			0
2100	8.25	20	19.2	0.71	87.8	9.7	135		124			0
2100	8.25	15	19.2	0.69	92.3	9.8	134		201			0
2100	8.25	15	27.4	0.89	87.7	8.9	109		128			0
2100	8.25	8	26.6	0.68	88.1	10.0	137		130			0
2100	8.25	20	19.5	0.80	91.4	9.4	121					0
2100	8.25	20	28.3	0.67	89.9	5.1	138		152			0
2100	8.25	15	27.7	0.82	89.5	7.5	118		167			0
2100	8.25	20	27.6	0.73	88.3	6.8	130		167			0
2100	8.25	8	19.0	0.69	90.0	11.3	135		138			0

(continued)

TABLE A-3 (continued)

Centri-fugal force x gravity	Pond set-ting	Scroll differ-ential RPM	Feed rate GPM	Feed consis-tency % solids	Solids recov-ery %	Cake consis-tency % solids	Uncondi-tioned sludge volume index	Condi-tioned sludge volume index	Unconditioned sludge specific resistence $x10^7$ sec^2/gm	Conditioned sludge specific resistence $x10^7$ sec^2/gm	Conditioning type	Condi-tioning amount lb/tn
2100	8.25	15	28.1	0.82	88.8	6.6	118		154			0
2100	8.37	6	19.3	0.77	89.9	10.2	127		133			0
2100	8.37	18	20.3	0.73	92.8	5.9						0
2100	8.37	12	20.1	0.69	92.2	5.9	140		90			0
2100	8.37	6	19.3	0.75	90.5	10.0	131		276			0
2100	8.37	18	31.3	0.66	93.2	3.1	150		171			0
2100	8.37	12	19.8	0.74	92.1	7.7	130					0
2100	8.37	6	27.3	0.61	91.1	6.5	158		151			0
2100	8.37	18	31.7	0.71	88.9	3.0	134		158			0
2100	8.37	12	31.7	0.76	93.7	3.3	127		159			0
2100	8.37	6	28.7	0.57	91.4	4.0	170		149			0
2100	8.37	18	20.8	0.77	94.7	5.5	127		157			0
2100	8.37	12	31.4	0.80	89.3	3.5	127		142			0
800	8.25	3	19.0	0.65	84.7	10.2	151		138			0
800	8.25	13	20.9	0.67	94.3	4.5	145		252			0
800	8.25	8	20.2	0.71	93.1	6.1	138		105			0
800	8.25	3	19.2	0.65	90.9	9.7	150		199			0
800	8.25	13	34.7	0.72	92.8	2.4	137		193			0
800	8.25	8	20.2	0.74	89.9	6.1	131		116			0
800	8.25	3	25.6	0.59	41.2	9.9	164		204			0
800	8.25	13	33.4	0.62	85.6	2.1	157		166			0
800	8.25	8	31.7	0.72	90.7	3.1	134		229			0
800	8.25	3	26.1	0.68	59.7	9.6	144		189			0
800	8.25	13	20.8	0.69	93.8	4.8	142		156			0
800	8.25	8	32.5	0.77	92.4	3.1	127		171			0
800	8.25	13	33.9	0.72	95.4	2.6		169	620	59	Hercufloc 844	3.1
800	8.25	13	63.5	0.69	97.3	1.6		169	323	57	Hercufloc 844	3.3
800	8.25	13	34.6	0.68	96.0	2.4		153	337	53	Hercufloc 844	5.6
800	8.25	13	57.3	0.71	97.4	1.9		137	905	70	Hercufloc 844	5.3
800	8.25	8	27.3	0.75	85.5	7.5		129	243	70	Hercufloc 844	7.1
800	8.25	8	47.4	0.73	98.4	2.9		124	722	74	Hercufloc 844	7.3
800	8.25	8	57.1	0.56	96.9	1.5		156	592	94	Hercufloc 844	4.0
800	8.25	8	28.6	0.69	92.8	5.1		134	143	90	Hercufloc 844	3.3
800	8.25	8	51.8	0.50	96.9	1.6		155	143	67	Hercufloc 844	7.6
800	8.25	8	27.9	0.69	94.7	6.3		131	167	70	Hercufloc 844	5.5

(continued)

TABLE A-3 (continued)

Centri-fugal force x gravity	Pond set-ting	Scroll differ-ential RPM	Feed rate GPM	Feed consis-tency % solids	Solids recov-ery %	Cake consis-tency % solids	Uncondi-tioned sludge volume index	Condi-tioned sludge volume index	Unconditioned sludge specific resistence x10^7 sec^2/gm	Conditioned sludge specific resistence x10^7 sec^2/gm	Conditioning type	Condi-tioning amount lb/tn
800	8.25	8	51.0	0.63	96.8	2.0	139	120		78	Hercufloc 844	8.5
800	8.25	8	27.7	0.72	96.1	7.1		129		108	Hercufloc 844	7.4
800	8.25	8	33.8	0.55	96.4	2.0		149		96	Hercufloc 844	4.1
800	8.25	8	32.0								Hercufloc 844	3.5
800	8.25	8	30.2	0.74	94.0	4.0		125		107	Hercufloc 844	5.1
800	8.25	8	46.6	0.64	96.4	2.7		146		74	Hercufloc 844	6.0
800	8.25	8	31.5	0.57	96.2	2.5					Betz 1260	4.0
800	8.25	8	28.3	0.75	98.7	6.4		78		47	Betz 1260	7.1
800	8.25	8	47.0	0.70	99.2	3.0		101		57	Betz 1260	7.6
800	8.25	8	48.2	0.64	97.9	2.5		79		29	Betz 1260	8.4
800	8.25	8	28.3	0.75	97.0	6.2		115	269	67	Betz 1260	3.1
800	8.25	8	29.3	0.67	97.7	4.4		86		41	Betz 1260	8.0
800	8.25	8	46.6	0.63	98.4	2.7		93		60	Betz 1260	3.6
800	8.25	8	41.2	0.73	98.9	5.8		91			Betz 1260	7.3
2100	8.25	6	26.7	0.61	85.8	8.1	162		106			0
2100	8.25	9	43.5	0.71	87.5	3.6	140		217			0
2100	8.25	15	29.9	0.72	91.0	4.0	138		124			0
2100	8.25	9	27.0	0.74	81.3	8.1	133		165			0
2100	8.25	6	39.8	0.76	67.5	5.4	130		102			0
2100	8.25	9	27.2	0.70	84.7	7.4	140		109			0
2100	8.25	15	47.4	0.75	89.8	2.8	132		192			0
2100	8.25	15	29.6	0.66	89.1	3.7	154		140			0
2100	8.25	15	45.9	0.74	89.1	3.0	134		124			0
2100	8.25	6	26.7	0.76	74.8	8.7	131		129			0
2100	8.25	9	42.4	0.74	82.3	4.0	134		107			0
2100	8.25	8	27.0	0.69		8.9				93	Betz 1260	3.3
2100	8.25	8	37.0	0.68	46.9	12.1				67	Betz 1260	3.3
2100	8.25	8	27.2	0.74	98.0	8.9				69	Betz 1260	5.2
2100	8.25	8	37.5	0.64		10.3				91	Betz 1260	6.0
2100	8.25	8	27.4	0.62	85.2	6.0				61	Betz 1260	8.6

(continued)

TABLE A-4. TAIT-ANDRITZ SDM DATA

Belt speed cm/min	Belt tension lb/linear in	Feed rate l/min	Feed consis- tency % solids	Conditioning type	Condi- tioning amount	Unconditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioned sludge specific resistance $\times 10^7$ sec^2/gm	Cake consis- tency % solids	Solids recovery %
137	40	24	0.95		0	62	62	22.5	78
274	40	28	0.83		0	60	60	21.9	40
548	40	26	0.81		0	60	60	11.8	26
274	60	25	0.81		0	60	60	20.2	29
274	20	24	0.93		0	63	63	20.4	59
137	20	24	0.92		0	64	64	19.6	80
274	40	29	0.87	Betz 1260	8.1	42	23	19.8	91
274	20	27	0.85	Betz 1260	8.6	40	19	17.7	85
274	60	29	0.94	Betz 1260	8.6	40	20	21.6	82
548	60	28	0.90	Betz 1260	8.6	36	23	18.0	42
548	20	28	0.90	Betz 1260	8.6	40	20	16.6	72
189	40	60	0.75	Betz 1260	6.6	51	28	19.5	98
189	40	55	0.77	Betz 1260	4.7	50	24	20.9	94
548	20	59	0.84		0	50	50	21.8	27
498	40	97	0.77	Betz 1260	6.8	50	24	19.4	86
498	40	103	0.84	Betz 1260	10.5	50	22	18.3	94
668	20	105	0.84	Betz 1260	11.2	50	24	17.2	90
551	20	164	0.84	Betz 1260	9.4	50	13	16.8	93
630	20	124	0.78	Betz 1260	11.8	51	11	16.2	86
269	20	74	0.72	Betz 1260	3.2	80	15	16.4	98
269	30	71	0.80	Betz 1260	4.3	120	22	18.3	90
668	30	52	0.87	Betz 1260	4.2	157	20	18.8	70
203	30	67	0.80	Betz 1260	3.7	150	20	18.1	90
274	50	68	0.75	Betz 1260	3.9	150	20	19.7	85
668	50	79	0.86	Betz 1260	3.4	150	20	19.9	53
668	10	69	0.85	Betz 1260	4.8	150	20	15.3	83
425	20	73	0.75	Betz 1260	6.2	225	46	13.7	92
425	40	73	0.78	Betz 1260	5.9	225	24	14.9	81
425	10	74	0.78	Betz 1260	5.8	225	15	13.4	94
452	10	72	0.78	Betz 1260	6.0	225	15	13.1	90
132	30	38	0.57	Betz 1260	8.7	541	50	16.4	87
132	10	36	0.51	Betz 1260	10.3	550	50	12.5	93
132	50	39	0.91	Betz 1260	3.5	550	19	17.3	85
312	50	75	0.98	Betz 1260	7.2	223	20	14.7	91
468	50	74	0.73	Betz 1260	9.8	200	35	16.0	76
275	30	70	0.79	Betz 1260	4.8	207	12	15.4	81

(continued)

TABLE A-4 (continued)

Belt speed cm/min	Belt tension lb/linear in	Feed rate l/min	Feed consis-tency % solids	Conditioning type	Condi-tioning amount	Unconditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioned sludge specific resistance $\times 10^7$ sec^2/gm	Cake consis-tency % solids	Solids recovery %
260	50	68	1.16	Betz 1260	6.0	100	28	27.2	93
493	30	78	1.04	Betz 1260	7.4	330	31	16.3	90
585	30	78	.98	Betz 1260	3.8	300	46	17.2	69
493	30	74	1.05	Betz 1260	6.4	300	22	15.4	88
425	30	72	1.22	Betz 1260	7.7	249	31	16.9	77
425	30	72	1.14	Betz 1260	10.4	250	21	16.1	87
292	30	70	1.18	Betz 1260	10.4	250	20	15.6	88
292	30	74	.81	Betz 1260	15.1	257	37	17.1	78
134	40	77	1.21	Betz 1260	24.4	250	1	23.2	95

TABLE A-5. SQUEEGEE DATA

Roll pressure lb/linear inch	Belt speed cm/sec	Feed rate ℓ/min	Feed consistency % solids	Cake consistency % solids	Solids recovery %	Unconditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioning type	Conditioning amount % of solids
4.2		10.0	0.7			240		$FeCl_3$	5.0
4.2	8.7	6.7	0.7	16.4	96		10	$FeCl_3$	5.0
4.2	8.7	6.7	0.6	15.6	99		57	$FeCl_3$	2.9
4.2	8.7	6.7	0.7	15.6	94		2	$FeCl_3$	10.0
4.2	8.7	8.6	0.7	16.7	91			$FeCl_3$	11.6
4.2	12.7	12.6	0.7	14.8	94		22	$FeCl_3$	5.2
4.2	8.4	6.6	0.7	15.1	97	234	25	$FeCl_3$	5.0
4.2	16.0	12.6	0.9	14.1	99		13	$FeCl_3$	5.0
1.2	16.0	12.6	0.9	15.4				$FeCl_3$	5.0
6.7	16.0	12.6	0.9	11.8		355		$FeCl_3$	5.0
6.7	8.4	5.0	0.9	12.5				$FeCl_3$	3.9
1.2	8.4	5.0	0.9	16.3				$FeCl_3$	3.9
4.2	8.4	5.0	0.9	15.1				$FeCl_3$	3.9
4.2	12.7	10.0	0.7	14.2	99		18	$FeCl_3$	5.0
1.2	12.7	10.0	0.7	16.1				$FeCl_3$	5.0
6.7	12.7	10.0	0.7	12.4				$FeCl_3$	5.0
6.7	8.4	10.0	0.7	12.0	96	74	14	$FeCl_3$	5.0
4.2	8.4	5.0	0.9	15.0	98		7	$FeCl_3$	3.8
1.2	8.4	5.0	0.9	16.6	96			$FeCl_3$	3.8
6.7	8.4	5.0	0.9	13.4	98			$FeCl_3$	3.8
6.7	8.4	5.0	0.9	12.7	99			$FeCl_3$	3.8
6.7	2.5	3.0	0.9	12.5	99			$FeCl_3$	3.9
6.7	2.5	3.0	0.9	13.7	98			$FeCl_3$	3.9
4.2	2.5	3.0	0.9	14.7	98			$FeCl_3$	3.9
1.2	2.5	3.0	0.9	16.4	95			$FeCl_3$	3.9
1.2	8.4	10.0	0.9	15.5	61	51		$FeCl_3$	3.9
6.7	2.5	2.5	0.8	12.6	98	48	28	$FeCl_3$	4.4
4.2	2.5	2.5	0.8	14.7	98			$FeCl_3$	4.4

(continued)

TABLE A-5 (continued)

Roll pressure lb/linear inch	Belt speed cm/sec	Feed rate l/min	Feed consistency % solids	Cake consistency % solids	Solids recovery %	Unconditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioning type	Conditioning amount % of solids
4.2	2.5	2.5	0.8	16.0	97			$FeCl_3$	4.4
1.2	2.5	2.5	0.8	17.0	90			$FeCl_3$	4.4
1.2	8.4	8.3	0.8	16.1	41			$FeCl_3$	4.4
4.2	5.6	4.0	0.9	15.7	97	39	22	$FeCl_3$	3.9
4.2	3.2	4.0	0.9	15.5	97			$FeCl_3$	3.9
4.2	4.0	4.0	0.9	15.5	96			$FeCl_3$	3.9
4.2	12.7	8.0	0.9	14.6	95			$FeCl_3$	3.9
4.2	5.6	8.0	0.9	13.8	96			$FeCl_3$	3.9
4.2	7.6	8.0	0.9	13.8	97			$FeCl_3$	3.9
4.2	15.2	12.0	0.9	13.7	95			$FeCl_3$	3.9
6.7	10.3	5.0	0.8	10.6	98	32			0.0
6.7	10.3	5.0	0.8	10.1				Percol 140	0.5
6.7	10.3	4.0	1.0	10.9	97		13	Hercufloc 859	0.5
6.7	5.1	5.0	1.0	11.1			27	Hercufloc 859	0.1
6.7	5.1	5.0	1.0	10.8		32			0.0
4.2	7.6	7.5	0.4	14.1	62	54		$FeCl_3$	7.9
4.2	10.1	10.0	0.6	15.9	92	45	5	$FeCl_3$	5.0
4.2	7.9	7.9	0.7	16.5	96			$FeCl_3$	5.0
4.2	3.3	3.4	0.7	16.8	97			$FeCl_3$	5.0
4.2	10.1	10.0	0.6	15.4	92	21	11	$FeCl_3$	5.8
4.2	10.1	13.0	0.6	14.7	94			$FeCl_3$	5.8
4.2	7.9	7.9	0.6	15.5	94			$FeCl_3$	5.8
4.2	10.1	10.0	0.6	14.7	96	30	15	$FeCl_3$	5.8
4.2	10.1	12.3	0.6	14.6	92			$FeCl_3$	5.8
4.2	3.3	3.4	0.6	16.8	94			$FeCl_3$	5.8
4.2	3.3	5.1	0.6	15.9	94			$FeCl_3$	5.8
4.2	3.3	4.1	0.6	16.5	98		8	$FeCl_3$	5.8
4.2	10.2	5.0	0.6	12.1	91			$FeCl_3$	5.8
4.2	10.1	10.0	0.6	14.9	96	50	18	$FeCl_3$	5.8

(continued)

TABLE A-5 (continued)

Roll pressure lb/linear inch	Belt speed cm/sec	Feed rate ℓ/min	Feed consistency % solids	Cake consistency % solids	Solids recovery %	Unconditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioned sludge specific resistance $\times 10^7$ sec^2/gm	Conditioning type	Conditioning amount % of solids
4.2	10.1	10.0	0.58	13.0	49		41	$FeCl_3$	2.9
4.2	10.1	10.0	0.59	13.5	96		17	$FeCl_3$	4.4
4.2	10.1	10.0	0.68	14.3	95		9	$FeCl_3$	5.0
4.2	10.1	10.0	0.65	14.6	95		2	$FeCl_3$	8.7
1.2	38.0	19.8	0.59	14.4	82			$FeCl_3$	10.1
1.2	31.7	10.9	0.67	12.9	77			$FeCl_3$	3.5
1.2	38.1	7.5	0.65	9.0	39				0
1.2	34.6	20.3	0.64	12.9	84			$FeCl_3$	6.7
1.2	6.7	6.0	0.65	18.8	92			$FeCl_3$	10.5
1.2	6.7	6.0	0.73	16.4	88			$FeCl_3$	4.7
1.2	6.7	6.0	0.72	12.9	69			$FeCl_3$	4.8
1.2	21.2	6.0	0.65	11.6	46			$FeCl_3$	5.5
1.2	21.8	6.0	0.76	11.9	83			$FeCl_3$	4.8
1.2	31.8	12.2	0.68	13.7	55			$FeCl_3$	5.1
1.2	34.6	12.2	0.62	9.6	80			$FeCl_3$	5.7
1.2	50.6	12.2	0.65	7.2				$FeCl_3$	5.4
1.2	25.4	12.2	0.80	6.8	29			$FeCl_3$	4.4

TABLE A-6 (Part I). PRESSURE FILTER DATA (Fresh Biological Solids)

Feed consistency % solids	Cycle time hr	Pressure atm	Precoat and conditioning % of sludge solids	Unconditioned specific resistance $\times 10^7$ sec^2/gm	Conditioned specific resistance $\times 10^7$ sec^2/gm	Bulk cake consistency - % solids	Solids recovery %	Weight of dewatered solids gm
1.2	2.0	12.0	DE Pre + 20% lime + 5% $FeCl_3$	59	17	45		790
1.2	3.0	12.0	DE Pre + 20% lime + 5% $FeCl_3$	72	26	47		670
1.4		12.0	DE Pre + 3.4# Betz 1260/ton run aborted					
1.6		7.5	DE Pre + 3.4# Betz 1260/ton + 8% lime run aborted					
0.8	3.0	7.5	DE Pre + 4.6# Betz 1260/ton	84	45	14	81	410
0.9	3.0	7.5	DE Pre + 4.6# Betz 1260/ton + 8% lime + 5% $FeCl_3$	172	52	35	99	320
2.4	3.0	7.5	DE Pre + 2# flyash/# sludge solids	198	57	37	98	560
0.9	2.0	7.5	lime mud Pre + .31# lime mud solids/# sludge solids	191	177			170
1.2	2.0	7.5	lime mud Pre + .75# lime mud solids/# sludge solids	160	95			270
1.7	1.0	5.1	DE Pre + 24% lime + 5.8% $FeCl_3$	54	5	33	98	810
0.7	2.0	7.5	DE Pre + 22.9% lime + 5.6% $FeCl_3$	22	8	35	99	470
1.6	2.0	7.5	DE Pre + 27.4% lime + 6.6% $FeCl_3$	26	8	37	99	1100
1.2	2.0	7.5	DE Pre + 1# broke solids/# sludge solids	21	42	35	99	540
0.8	2.0	7.5	DE Pre + .30# bark fines/# sludge solids	160	87	27	98	250
1.0	2.0	7.5	DE Pre + .5# broke solids/# sludge solids	33	16	30	99	330

(continued)

TABLE A-6 (Part I) (continued)

Feed consistency % solids	Cycle time hr	Pressure atm	Precoat and conditioning % of sludge solids	Unconditioned specific resistance $\times 10^7$ sec^2/gm	Conditioned specific resistance $\times 10^7$ sec^2/gm	Bulk cake consistency - % solids	Solids recovery %	Weight of dewatered solids gm
0.9	3.0	7.5	20.9% lime + 5.2% $FeCl_3$	50	15	33	99	510
1.0	2.2	7.5	20% lime + 5# Betz 1260/ ton	49	292	15	98	230
0.9	1.3	7.5	21.3% lime + 6.3# Betz 1260/ton	48	317			
1.1	2.0	7.6	18.8% lime + 4.7 $FeCl_3$	53	33	30		460
0.8	2.0	13.5	DE Pre + 29.3% lime + 7.3% $FeCl_3$	40	17	28		390
0.7	2.0	10.6	DE Pre + 24.5% lime + 6.1% $FeCl_3$	36	21	27		310
0.8	2.0	6.8	DE Pre + 24.5% lime + 6.1% $FeCl_3$	52	26	29		340
1.3	3.3	13.5	DE Pre + 31.9% lime + 7.9% $FeCl_3$	68	13	42		800
1.2	1.2	6.8	DE Pre + 38.5% lime + 9.6% $FeCl_3$	62	7	38		710
1.7	2.4	6.8	DE Pre + 34.1% lime + 8.5% $FeCl_3$	77	10	39		700
0.9	2.3	13.5	33% lime + 8.2% $FeCl_3$	108	13	35		520
0.9	2.3	10.5	100% lime + 25% $FeCl_3$			35		650

TABLE A-6 (Part II). PRESSURE FILTER DATA
(Degraded Biological Solids)

Feed consistency % solids	Cycle time hr	Pressure atm	Precoat and conditioning % of sludge solids	Unconditioned specific resistance $\times 10^7$ sec^2/gm	Conditioned specific resistance $\times 10^7$ sec^2/gm	Bulk cake consistency - % solids	Solids recovery %	Weight of dewatered solids gm
1.2	3.0	7.5	DE Pre + 45% lime + 4.5% $FeCl_3$	1060	20	41	99	640
1.1	3.0	7.5	DE Pre + 45% lime + 4.5% $FeCl_3$	1060	14	40	98	830
1.6	2.0	7.5	DE Pre + 29% lime	243	126	37	92	420
1.4	2.0	7.5	DE Pre + 33% lime + 3.3% $FeCl_3$	396	19	40	98	640
0.7	3.0	7.5	DE Pre + 32% lime + 3.1% $FeCl_3$	185	63	32	99	200
1.6	2.0	7.5	DE Pre + 11.5% lime + 2.9% $FeCl_3$	99	46	29	99	320
1.3	2.0	13.5	DE Pre + 15.3% lime + 3.7% $FeCl_3$	74	48	23	99	260
1.4	2.0	5.1	DE Pre + 15.9% lime + 3.8% $FeCl_3$	73	23	27	99	280
1.4	2.0	5.1	DE Pre + 28.6% lime + 7% $FeCl_3$	137	7	33	99	680
1.6	2.0	7.5	DE Pre + 25% lime + 6.1% $FeCl_3$	71	8	37	99	680
1.4	2.0	13.5	DE Pre + 26% lime + 6.3% $FeCl_3$	68	9	39	99	590
1.3	1.0	13.5	DE Pre + 34% lime + 8.3% $FeCl_3$	70	4	36	99	540
1.3	1.0	7.5	DE Pre + 35% lime + 8.4% $FeCl_3$	76	2	35	99	590
2.2	2.0	7.5	DE Pre + 22.5% lime + 5.6% $FeCl_3$	78	8	36	99	720
0.6	2.0	7.5	DE Pre + 22.4% lime + 5.6% $FeCl_3$	37	20	32	99	250
2.7	1.4	7.5	DE Pre + 15.7% lime + 3.9% $FeCl_3$	94	15	37	99	650
	2.8	7.5	DE Pre + 15% lime + 4% $FeCl_3$	94	15	38	99	700

(continued)

TABLE A-6 (Part II) (continued)

Feed consistency % solids	Cycle time hr	Pressure atm	Precoat and conditioning % of sludge solids	Unconditioned specific resistance $\times 10^7$ sec^2/gm	Conditioned specific resistance $\times 10^7$ sec^2/gm	Bulk cake consistency - % solids	Solids recovery %	Weight of dewatered solids gm
0.9	2.0	6.8	DE Pre + 19.9% lime + 5% $FeCl_3$	393	24	27		320
1.0	2.0	10.5	DE Pre + 20% lime + 5% $FeCl_3$	324	23	28		380
1.0	2.0	13.5	DE Pre + 19.7% lime + 4.9% $FeCl_3$	246	29	27		350
1.1	1.6	6.8	32.3% lime + 8% $FeCl_3$	198	16	32		631
0.9	3.3	13.5	30.8% lime + 7.7% $FeCl_3$	289	15	37		540
1.4	1.3	10.5	.6# lime mud solids/# sludge solids + 4# Percol 140/ton	214	46			
1.7	3.5	6.8	1# flyash/# sludge solids + 4# Percol 140/ton	152	40			510
0.8	2.9	10.5	45.2% lime + 11.2% $FeCl_3$	110	20	34		470
0.9	3.1	10.5	44.7% lime + 11.1 $FeCl_3$	96	16	35		530
11.2	1.2	11.0	30% lime + 7.5% $FeCl_3$			41		880

APPENDIX B

PERFORMANCE LEVELS USED TO CALCULATE EQUIPMENT REQUIREMENTS

ULTRAFILTER

Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency

Average flux rate: 80 gal/ft^2/day
Equipment requirements: 270 modules (2540 ft^2)
Conditioning: none
Power requirements: for 1 pump (as utilized in the pilot study) for every third stage connected HP is about 300 HP
Cake Consistency: 6 percent solids
Solids recovery: 100 percent

Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency

Average flux rate: 50 gal/ft^2/day
Equipment requirements: 160 modules (1500 ft^2)
Conditioning: none
Power requirements: under same circumstances as above connected HP is about 250 HP
Cake consistency: 6 percent solids
Solids recovery: 100 percent

SHARPLES P3000 BD Centrifuge

Basis: 10 tpd of Waste Activated Solids at 1 Percent Consistency

Loading: 95 gpm/unit
Equipment requirements: 2 P5400-BD centrifuges (20 percent additional capacity included)
Conditioning: none
Power requirements: 320 connected HP, 120 running HP
Cake consistency: 8 to 10 percent solids
Solids capture: 90 percent

Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency

Insufficient data

SQUEEGEE

<u>Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency</u>

Loading: 2.6 gpm/ft of belt width
Equipment requirements: units totaling 75 feet of belt width (includes 20 percent additional capacity)
Conditioning: 5 percent $FeCl_3$
Power requirements: estimated to be approximately 60 connected HP
Cake consistency: 15 to 17 percent solids
Solids recovery: 90 percent

<u>Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency</u>

Insufficient data

PERMUTIT DCG-MRP

<u>Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency</u>

Loading: 30 gpm/DCG 200 and 1 MRP per 2 DCG 200's (2 DCG 100's = DCG 200)
Equipment requirements: 7 DCG 200's and 4 MRP's (includes (20 percent addiitonal capacity)
Conditioning requirements: 5 to 10 pounds Betz 1260/ton
Power requirements: 10 to 12 connected HP
Cake consistency: 15 to 17 percent solids
Solids recovery: 90 percent +

<u>Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency</u>

Insufficient data

PRECOAT VACUUM FILTER

<u>Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency</u>

Loading rate: 1 $lb/ft^2/hr$ (1/2 RPM, 40 percent submergence)
Equipment requirements (allowing 20 percent additional capacity): two 500-ft^2 filters
Precoat consumption: at 5 percent of sludge solids = 1000 lb/day
Power requirements: 132 connected HP during filtration, an additional 63 during precoating
Cake consistency: 25 to 30 percent solids
Solids recovery: 99 percent +

PRECOAT VACUUM FILTER (cont.)

Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency

Loading rate: 2 $lb/ft^2/hr$ (1/2 RPM, 40 percent submergence)
Equipment requirements: one 500-ft^2 filter (includes 20 percent additional capacity)
Precoat consumption: at 2.5 percent of sludge solids = 500 lb/day
Power requirements: 66 connected HP during filtration, an additional 63 connected HP during precoating
Cake consistency: 25-30 percent solids
Solids recovery: 99 percent +

TAIT-ANDRITZ SDM

Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency

Loading rate: 15 gpm/ft of belt width
Equipment requirement (allowing 20 percent additional capacity): two 80-inch SDM;s
Conditioning requirements: 8 pounds Betz 1260/ton
Power requirements: 10 connected HP
Cake consistency: 17 to 20 percent solids
Solids recovery: 90 percent +

Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency

Insufficient data

NETZSCH PRESSURE FILTER

Basis: 10 tpd Waste Activated Solids at 1 Percent Consistency

Loading rate: 0.30 $lb/ft^2/hr$ with 1-inch thick cake or 1.31 lb/gal/hr (total solids basis)
Equipment requirements: 1 press containing 80 5- by 5-foot plates
Conditioning: 35-40 percent lime + 8-10 percent $FeCl_3$
Power requirements: 20 HP average over cycle
Cake consistency: 32-38 percent bulk (22 to 28 percent sludge solids basis)
Solids recovery: 99 percent

Basis: 10 tpd Waste Activated Solids at 2 Percent Consistency

Loading rate: 0.39 $lb/ft^2/hr$ with 1-inch thick cake or 1.25 lb/gal/hr (total solids basis)
Equipment requirements: 1 press containing 90 5- by 5-foot plates (includes 20 percent extra capacity)
Conditioning: 25 to 30 percent lime + 7 percent $FeCl_3$
Power requirements: 20 HP average over cycle
Cake consistency: 38 to 40 percent
Solids recovery: 99 percent

APPENDIX C

MANUFACTURERS OF PRESSURE FILTERS

Edwards-Jones — Phone: 416-869-1463
William R. Perrin
530 King Street East
Toronto, Ontario   M5A1M1

Netzsch — Phone: 215-935-9201
P-K Associates
Box 701
Valley Forge, Pennsylvania   19481

Passevant Corporation — Phone: 205-853-6290
Carson Road
Birmingham, Alabama   35201

Shriver-Johnson — Phone: 201-484-2500
T. Shriver and Company, Inc.
808 Hamilton Street
Harrison, New Jersey   07920

APPENDIX D

MANUFACTURERS OF PRECOAT VACUUM FILTERS

Ametek, Inc.
76 Thomas Street
East Moline, Illinois 61244

Dorr-Oliver, Inc.
77 Havermeyer Lane
Stamford, Connecticut 06904

Envirotech Corporation
669 West Second South
P.O. Box 300
Salt Lake City, Utah 84110

Komline-Sanderson Engineering Corporation
100 Holland Avenue
Peapack, New Jersey 07977

APPENDIX E

MANUFACTURERS OF FILTER BELT PRESSES

Dravo One Oliver Plaza Pittsburgh, Pennsylvania 15222	Phone: 412-566-3492
Environmental Machine and Mfg. Company 1151 N.W. 6th Street Gainesville, Florida 32601	Phone: 904-375-2450
Passevant Corporation Carson Road Birmingham, Alabama 32501	Phone: 205-853-6290
Permutit Company, Inc. 8700 North Waukegan Road Morton Grove, Illinois 60053	Phone: 312-967-5071
Ralph B. Carter Company 6659 North Avondale Chicago, Illinois 60631	Phone: 312-631-0244
Smith and Loveless Division Econdyne Corporation Lenexa, Kansas 66215	Phone: 913-888-5201
Tait-Andritz Box 1138 Lubbock, Texas 70408	Phone: 806-747-8666

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-78-014	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Pilot Investigations of Secondary Sludge Dewatering Alternatives		5. REPORT DATE February 1978 issuing date 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Reid A. Miner, Duane W. Marshall, Isaiah Gellman *		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS National Council of the Paper Industry for Air and Stream Improvement, Inc. Western Michigan University (Central-Lake States Reg.Ctr.) Kalamazoo, Michigan 49008		10. PROGRAM ELEMENT NO. 1BB037 11. CONTRACT/GRANT NO. R-804019-01-0
12. SPONSORING AGENCY NAME AND ADDRESS Industrial Environmental Research Laboratory--Cin..OH Office of Research & Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268		13. TYPE OF REPORT AND PERIOD COVERED Final 11/1/75 to 10/31/76 14. SPONSORING AGENCY CODE EPA/600/12

15. SUPPLEMENTARY NOTES

*Isaiah Gellman is with the National Council of the Paper Industry for Air and Stream Improvement, Inc, New York, New York 10016

16. ABSTRACT

A pilot investigation of biological sludge thickening and dewatering alternatives, including pressure filtration, precoat vacuum filtration, filter belt pressing, capillary suction, dewatering, gravity filtration, centrifugation, and ultrafiltration has been conducted on waste activated sludge resulting from the treatment of wastewater from an integrated bleached kraft-fine paper mill. Based upon a criterion of attainable cake consistency, three levels of performance are indicated: (1) pressure filtration and precoat vacuum filtration generating the driest cakes, (2) filter belt pressing yielding intermediate cake consistencies, and (3) gravity filtration, centrifugation, and ultrafiltration resulting in relatively low cake consistencies. Performance was found to be severely affected by changes in feed sludge consistency, the amount of sludge conditioning, and the sludge's specific resistance to filtration. The type and amount of sludge conditioning required was extremely variable, depending upon the dewatering technique employed, the level of performance expected of it, and the consistency and nature of the sludge dewatered.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pulp mills, Paper mills, Sludge disposal, Vacuum filtration, Pressure filters, Dewatering, Centrifuging	Sludge dewatering, Precoat vacuum filtration Sludge conditioning	50 B

18. DISTRIBUTION STATEMENT Release to public	19. SECURITY CLASS (This Report) UNCLASSIFIED	21. NO. OF PAGES 127
	20. SECURITY CLASS (This page) UNCLASSIFIED	22. PRICE

EPA Form 2220-1 (9-73)

☆ U. S. GOVERNMENT PRINTING OFFICE: 1978-757-140/6696 Region No. 5-11