SEMINAR ON PHOSPHORUS REMOVAL
MAY 1-2, 1968
WALDORF ROOM -- CONRAD HILTON HOTEL
CHICAGO, ILLINOIS

May 1		Speaker
9:30 a.m.	Welcome & Introductory Remarks	H. W. Poston
	Enforcement Conference Recommendations	M. Stein
	Seminar Objectives	Dr. L. W. Weinberger

1st TECHNICAL SESSION
Moderator - Dr. D. G. Stephan

10:00 a.m.	Forms in Measurement of Phosphorus	Dr. R. B. Dean
10:30 a.m.	Present & Projected Phosphorus Removal in Conventional Treatment	Dr. R. L. Bunch
11:15 a.m.	Break Period	
11:30 a.m.	Phosphorus Removal by Mineral Additions to Primary - Secondary Processes	E. F. Barth
12:15 p.m.	Lunch	

2nd TECHNICAL SESSION
Moderator - F. M. Middleton

2:00 p.m.	Phosphorus Removal by Tertiary Treatment with Lime & Alum	Dr. C. A. Brunner
2:45 p.m.	Reuse & Disposal of Lime Sludges	Dr. R. B. Dean
3:30 p.m.	Break Period	
3:45 p.m.	Handling, Recovery & Disposal of Alum Sludges	Dr. J. B. Farrell

May 2	3rd TECHNICAL SESSION	
	Moderator - A. C. Printz, Jr.	
9:00 a.m.	Alternative Methods for Phosphorus Removal	J. M. Cohen
9:45 a.m.	Project Costs for Phosphorus Removal & Sludge Disposal	R. Smith
10:30 a.m.	Break Period	
10:45 a.m.	Removal of Nitrogen from Waste Waters	
	Part I - Biological	E. F. Barth
	Part II - Physical-Chemical	Dr. J. B. Farrell
11:30 a.m.	FWPCA - Supported Phosphorus Removal Projects	Dr. D. G. Stephan
12:00 noon	Announcements	Dr. J. I. Bregman
12:15 p.m.	Closing Remarks	H. W. Poston

ENVIRONMENTAL PROTECTION AGENCY

ENVIRONMENTAL PROTECTION AGENCY

Forms and Measurements of Phosphorus   R. B. Dean

Phosphoric Anhydride P_4O_{10} (usually calculated as P_2O_5)

used in report of fertilizer ... *10 parts P_2O_5*

Orthophosphates Found in Water

Phosphoric acid	H_3PO_4	below pH 2.2
Dihydrogen phosphate ion (monovalent)	$H_2PO_4^-$	from pH 2.2 to 7.2
Monohydrogen phosphate ion (divalent)	$HPO_4^=$	from pH 7.2 to 12.4
Phosphate ion (trivalent)	$PO_4^=$	above pH 12.4
Dicalcium phosphate	$CaHPO_4$	from pH 6 to 8

(Tricalcium phosphate, $Ca_3(PO_4)_2$, does not form in water)

Hydroxyapatite	$Ca_5OH(PO_4)_3$	above pH 7-8
Fluorapatite	$Ca_5F(PO_4)_3$	above pH 7

Aluminum hydroxide adsorbs $HPO_4^=$ from pH 5 to 10

Aluminate ion $Al(OH)_4^-$ forms above pH 10

Ferric hydroxide adsorbs $HPO_4^=$ above pH 4

Magnesium hydroxide flocculates colloidal phosphates above pH 10-11

Polyphosphates

All polyphosphates hydrolyze slowly to orthophosphates

Sodium Trimetaphosphate	$Na_3(PO_3)_3$	-- a ring compound
Sodium Polymetaphosphate	$(NaPO_3)_n$	-- a long chain

(there is no justification for the name "hexametaphosphate")

Sodium Pyrophosphate	$Na_4P_2O_7$
Sodium Tri Polyphosphate	$Na_5P_3O_{10}$

Organic Phosphates

Esters and anhydrides	*ADP ATP*	$ROPO_3^=$, $ROPO_3PO_3^=$
Phosphagens		$RNPO_3^=$

Reference

(1) Van Wazer, J. R. Phosphorus and Its Compounds. Interscience 1958.

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<u>Forms and Measurements of Phosphorus</u>   R. B. Dean

Tentative FWPCA Definitions for Phosphorus Forms in Water

All expressed as mg/l of P

<u>Method</u>

I.   P, Soluble (filterable)   A + B + C

II.   P, Total   B + C

III.   P, Soluble, Ortho   A + C

IV.   P, Total, Ortho (see C below)   C

Methods for Determining Forms of P

A.   Soluble: Filter through 0.45 micron (bacteriological) membrane filter.

      Soluble or filterable phosphorus consists of truly soluble
      ions, colloidal precipitates of insoluble phosphates,
      fragments of cells and organic and inorganic phosphates
      (and polyphosphates) adsorbed on natural colloids.

      Insoluble phosphorus equals Total minus Soluble and consists
      of phosphorus in cells and multicellular matter together
      with clay and other minerals carrying adsorbed phosphates.

B.   Total: Convert to Ortho by digestion with acid and persulfate.
      M. E. Gales, Jr., E. C. Julian and R. C. Kroner, JAWWA $\underline{58(10)}$,
      1363-8 (1966).

      Digestion converts all polyphosphates and organic phosphates
      to orthophosphoric acid and destroys organic turbidity and
      color.

      Acid hydrolysis liberates most inorganic phosphates and some
      organic phosphates.

C.   Ortho: Mixed reagent test using ascorbic acid.
      J. Murphy and J. Riley, Anal. Chem. Acta. $\underline{27}$, 31 (1962).

      or perhaps Stannous Chloride Method, Standard Methods for the
      Examination of Water and Wastewater, 12th edition (1965) in
      fresh water samples.

      Phosphoric acid is converted to phosphomolybdic acid and
      reduced to a blue color. Corrections must be made for color
      and turbidity of the sample when determining P, Total, Ortho. (IV)

Forms and Measurements of Phosphorus

R. B. Dean

Conversion Factors

In water analyses all measurements should be reported as mg/l of P.

To convert X to P Multiply by:	X	To convert P to X Multiply by:
1.00	P	1.00
.451	P_2O_5	2.29
.327	PO_4	3.06
.316	H_3PO_4	3.16
.200	$Ca_3(PO_4)_2$	5.00
.181	$Ca_5OH(PO_4)_3$	5.51

May 1, 1968

SEMINAR ON PHOSPHORUS REMOVAL
CHICAGO, ILLINOIS
MAY 1 AND 2, 1968

REMOVAL OF PHOSPHATE FROM MUNICIPAL SECONDARY EFFLUENT
BY LIME TREATMENT

Carl A. Brunner

A 75-gpm lime treating pilot plant is being operated at the Lebanon,
Ohio Sewage Treatment Plant for clarification of secondary effluent. During
operation of the pilot system, data have been obtained on removal of phos-
phate both in the clarifier and on the filters following the clarifier.

The clarifier is a single-stage, upflow unit with internal recircu-
lation of sludge. This equipment is shown in Figure 1. The upflow rate
in the settling section of the unit is 1 gpm/ft^2. Detention time in the
clarifier is 110 minutes. High-calcium hydrated lime is used for clarifi-
cation. It is added as a slurry at a rate sufficient to maintain a pre-
determined pH in the clarifier. The slurry contacts incoming secondary
effluent in the mixing zone. The mixture then flows into the flocculation
zone. Water is moved up through the mixing zone at a high enough rate to
draw some of the sludge off the bottom of the clarifier into that zone and
out into the flocculating zone. The purpose of the sludge recirculation is
to increase sludge particle size and to hasten precipitation of calcium
carbonate and other inorganic materials. The water flows downward in the
flocculating zone and either enters the settler or returns to the mixing
zone. Sludge deposits on the bottom of the clarifier and is moved by
tangential pulses of water to the collection ring. The water pulse system
is not shown in Figure 1. It is possible to concentrate the sludge to 10
percent solids in the clarifier, but clogging problems may result. Usually,
the sludge is removed as a 2 or 3-percent slurry and is gravity thickened
to 10 percent solids in a separate vessel.

Following the clarifier are two dual-media filters that operate in
parallel. The media beds consist of 6 in. of 0.46-mm sand overlaid with
18 in. of 0.75-mm anthracite coal. Water rate through the filters is
2 gpm/ft^2. Filters are backwashed when the pressure drop exceeds 7 in. Hg.
Filter runs usually exceed 48 hrs and are often much longer.

The secondary effluent that is being lime clarified is a hard water
with high alkalinity and high ionic strength. When there is no storm-water
dilution of the sewage, the calcium content is about 100 mg/l as Ca,
magnesium content is about 30 mg/l as Mg, and alkalinity is about 400 mg/l
as $CaCO_3$. Lime treatment of this water gives a rapidly-settling sludge.
The amount of insoluble material that is carried out of the settler is
small, averaging about 15 mg/l. This material is largely calcium carbonate.
It is effectively removed by the dual-media filters.

Phosphate removal in the clarifier alone and over the whole clarification system has been determined as a function of clarifier pH. These results are shown in Figures 2 and 3. The average phosphate concentration of the secondary effluent during this study was 30 mg/l as PO_4. Phosphate removal improves as operating pH is increased. This effect is most clearly shown when the water has been filtered. (It is not intended to discuss here the forms in which the residual phosphate may exist. Because of the very low turbidity of the filtered water, however, it appears that the phosphate is in true solution.) The data in Figure 3 give a good indication of the degree of phosphate removal that can be obtained by clarification of a hard, high alkalinity water using single-stage treatment. An average phosphate concentration of less than 1 mg/l is obtainable at a pH of 10. For this particular water, 90-percent phosphate removal is obtainable down to a pH of 9.5. Eighty-percent removal occurs at a pH of about 9.0.

Where phosphate removal is the primary purpose of lime treatment, elimination of filtration would be desirable because of the resulting cost savings. The residual phosphate concentration would be expected to increase because of the presence of phosphate-containing suspended solids. The data of Figure 2 show, however, that good phosphate removal is still possible. Ninety-percent removal is generally obtainable down to a pH of 9.7; 80-percent removal down to slightly lower pH. If improved removal were required, a lower rate of rise could be used in the settler. This would decrease the carry-over of suspended solids and any phosphate contained in these solids. The cost of a larger settler would be substantially less than the cost of filters.

An important factor in the economics of lime treatment is the amount of lime required. Lime dose depends upon the mineral composition of the water. One of the most important components is the alkalinity. The clarifier pH is also important because the reactions that occur during lime treatment are pH dependent. Table 1 gives lime doses that were measured for a range of alkalinities and pH values. Because other constituents in wastewater may have a significant effect on lime dose, the values in Table 1 must be considered very approximate. The importance of alkalinity on dose is obvious, however. For the Lebanon plant it is estimated that an annual average of about 250 mg/l of hydrated lime would be required to obtain 80-percent removal of phosphate. A lower lime requirement is likely at many plants because of lower alkalinity.
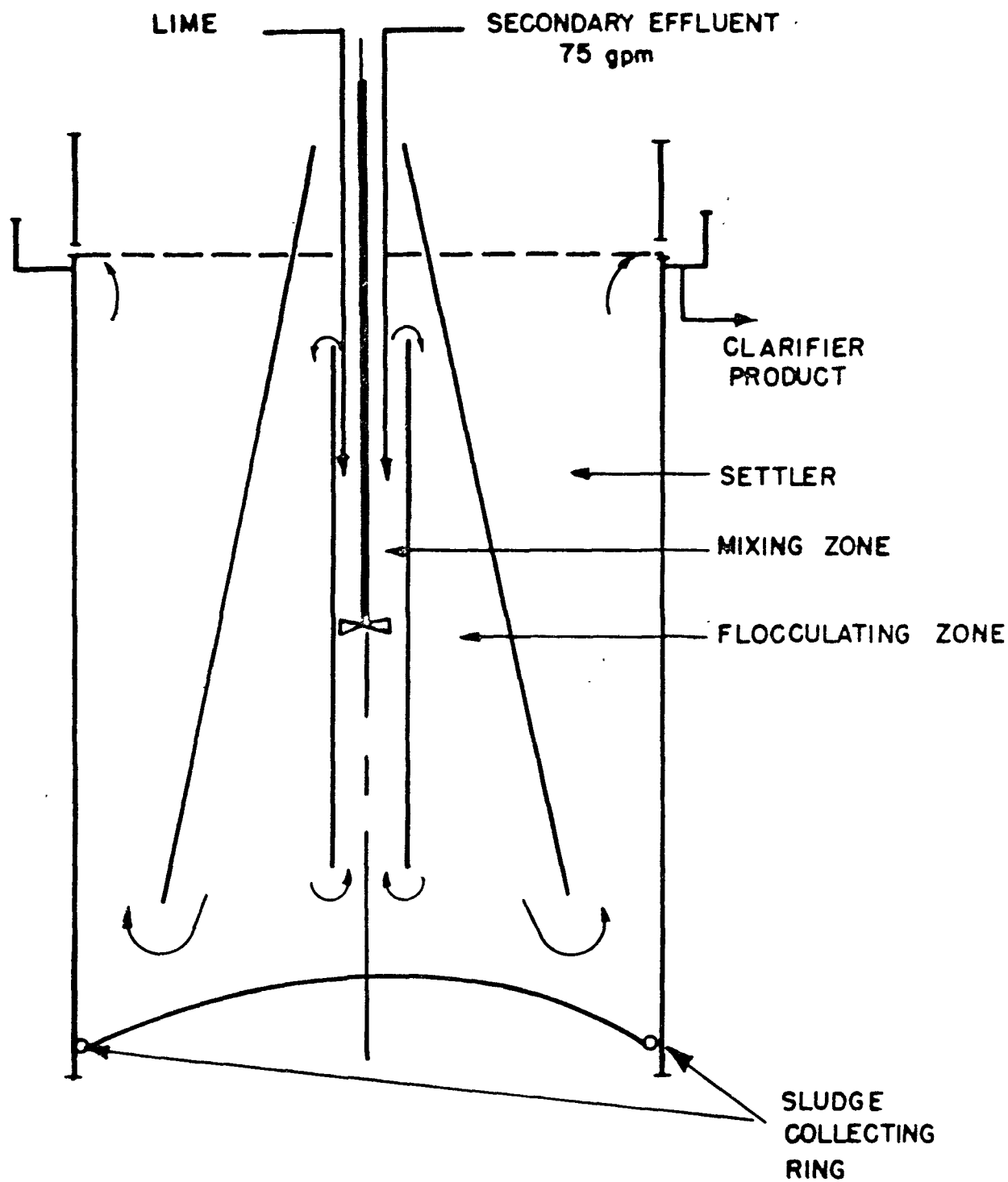
LIME   SECONDARY EFFLUENT
              75 gpm

CLARIFIER
PRODUCT

SETTLER

MIXING ZONE

FLOCCULATING ZONE

SLUDGE
COLLECTING
RING

FIGURE I. LIME CLARIFIER

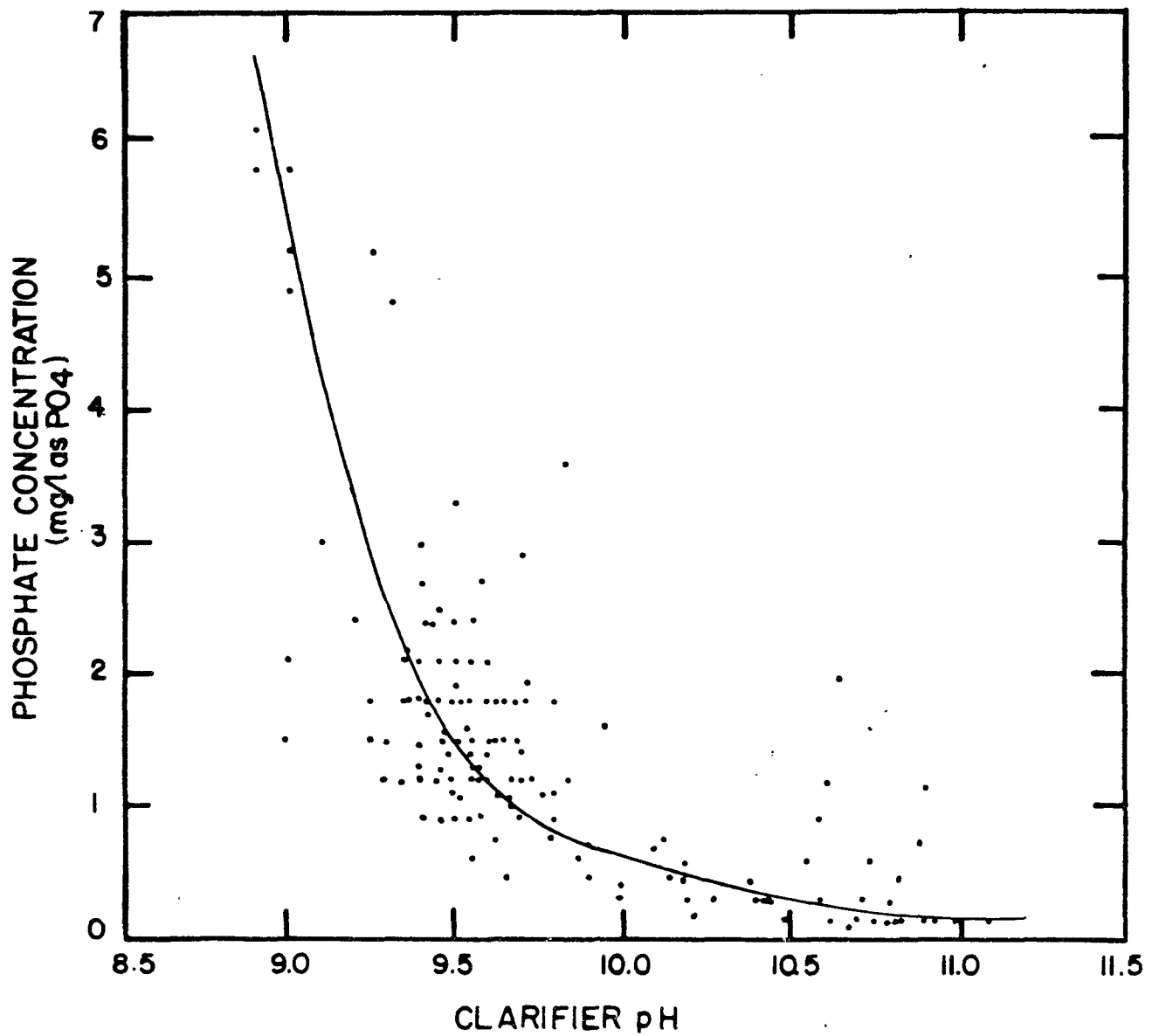FIGURE 2. EFFECT OF pH ON PHOSPHATE CONCENTRATION OF EFFLUENT FROM LIME CLARIFIER

FIGURE 3. EFFECT OF pH ON PHOSPHATE CONCENTRATION OF
EFFLUENT FROM FILTERS FOLLOWING
LIME CLARIFIER

TABLE I. LIME REQUIREMENTS

FEED WATER ALKALINITY (mg/l as $CaCO_3$)	CLARIFIER pH	APPROXIMATE LIME DOSE (mg/l of $Ca(OH)_2$)
300	9.5	200
300	10.5	350
400	9.5	300
400	10.5	500

MINERAL CONTROLLED PHOSPHORUS REMOVAL*

E. F. Barth and M. B. Ettinger, Research Chemists,
Federal Water Pollution Control Administration,
Cincinnati Research Laboratory, Cincinnati, Ohio

When considering the removal of phosphorus by biological treatment processes, it becomes evident that the cycle of phosphorus in a biological system is different from the usual wastewater components such as carbon and nitrogen. This is illustrated in Table I. Phosphorus removal is probably a combination of cellular growth and inorganic solubility products. This may account for the wide variation in phosphorus removals reported from various municipal treatment plants. The mineral composition of the wastewaters or characteristic industrial wastes may influence the phosphorus removal.

Many investigators have studied the problem of phosphorus removal. Figure 1 indicates some of the approaches. It will be noted that the approaches can be divided into two groups. One group attempts control by biological means; the other group adjusts the mineral composition of the biological effluent to precipitate phosphorus.

Our approach to the problem was to blend a chemical precipitation with the active biological solids. We did this by introducing, directly into the aeration chamber, mineral salts that were known to form slightly soluble phosphorus compounds. The work was done on a 100-gallon-per-day continuous flow pilot plant, operated as conventional activated sludge.

The mineral addition tried and the phosphorus removals obtained are given in Table II. During the periods when iron and aluminum were added to the aerator, turbid effluents were obtained. The pH in the aerator dropped to about 6.2 because of hydrolysis of the metal salt producing three hydrogen ions for each ion of metal added. This was corrected, as shown in the last entry in Table II, by introducing calcium with the aluminum to control pH.

In order to reduce the extraneous sulfate, chloride, or calcium ions introduced with this approach to phosphorus removal, it was decided to investigate the use of sodium aluminate. This material will introduce only a small extraneous increment of sodium and will automatically compensate for pH control in

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*A more complete publication will appear in the Journal of the Water Pollution Control Federation.

the aerator. Excess hydroxyl ions will be converted to bicarbonate by the carbon dioxide produced during aeration.

Table III gives the efficiency of phosphorus removal when introducing sodium aluminate directly to the aerator. The amounts of aluminum and phosphorus were varied to determine the dosage necessary for good removal. It can be seen from Table III that when the aluminum-phosphorus ratio is about 1:1 good removals are obtained.

At this point in our studies we cannot define the relative contribution of simple chemical precipitation or sorptive properties of the activated sludge mass to overall phosphorus removal. The dosage of precipitant needed is much less than that needed for a separate unit operation without biological solids. This indicates that the large surface area of the biological floc is helpful.

The aluminum does not in any way interfere with biological nitrification or carbon or solids removal.

The sludge produced by the mineral supplement and the microbial floc are intimately associated and the existing final settler serves as the liquid-solids separation device, saving an additional unit operation. The mixed sludge produced has better settling characteristics than either a biological sludge or aluminum hydroxide floc alone.

TABLE I

Difference Between Phosphorus and Other Wastewater

Components During Biological Treatment

Materials that can show a <u>net loss</u> through treatment.
Not conserved in process streams.

		Aerobic	Anaerobic
BOD COD }	C,H	CO_2,H_2O	$\overset{\bullet}{CH_4H_2}$, CO_2
Nitrogen		NO_2,NO_3,NH_3	$NH_3,\overset{\bullet}{N_2}$
Sulfur		SO_4	H_2S,S
Solids		All above	All above

Phosphorus is conserved. <u>No net loss</u> through treatment,
should be able to account for every bit of phosphorus in
process streams. Phosphorus enters in highest oxidized
form, no common biological systems reduce phosphate.

	Aerobic	Anaerobic
Phosphorus	Organic P $\rightleftharpoons$ Inorganic P	Organic P $\rightleftharpoons$ Inorganic P

Phosphorus removal is probably a combination of cellular growth and
inorganic solubility.

TABLE II

Results of Screening for Mineral Supplement
to Activated Sludge Process
Direct Dosage to Aerator
No Supernatant Recycle, No Primary Settling

Mineral Addition	Introduced as	To Form	Overall Removal
None (base line)	-	-	40%
Ca, 150 mg/l	CaO	Hydroxyapatite	64%
Ca, 150 mg/l plus F, 6 mg/l	CaO NaF	Apatite	75%
Mg, 20 mg/l	$MgSO_4$	$MgNH_4PO_4$	50%
Fe, 15 mg/l	$FeCl_3$	$FePO_4$	75%*
Al, 20 mg/l	$Al_2(SO_4)_3$	$AlPO_4$	70%*
Al, 30 mg/l plus Ca, 20 mg/l	$Al_2(SO_4)_3$ CaO	$AlPO_4$	90%

* Turbid effluents

TABLE III

Relation of Aluminum to Phosphorus

Direct Dosage of NaAl(OH)$_4$ to Aerator

Al^{+3}, mg/l	Influent Phosphorus, mg/l	Effluent Phosphorus, mg/l	Percent Removal (mass balance)
5	3	0.2	94
5	8	2.2	75
10	13	1.3	90
10	10	0.2	98

FIGURE I

Phosphorus Removal by Wastewater

Treatment (Excluding Lagoons and Percolation)

PHOSPHORUS REMOVAL DEPENDENT UPON:

Mineral Composition   Interaction   Biological Synthesis

Addition of Supplement
to Primary Effluent

Guggenheim (1)*

E. A. Thomas (2)

Separate Unit Process
on Biological Effluent

Rohlich (3)

Culp (4)

Yee (5)

Stumm and
Tenney (6)

Control of Conventional
Operational Parameters

Levin and Shapiro (7)

Vacher et al. (8)

Primary Settling Efliminated
to Increase C:N:P

Finstein (9)

Algae Culture

Gates (11)

Eliminate Digester
Supernatant Feedback
Westberg (10)

*( ) indicates literature reference.

PHOSPHORUS REMOVAL IN CONVENTIONAL TREATMENT

Robert L. Bunch

TREATMENT CLASSIFICATION BY POPULATION FOR YEAR 1962

Population	Type of treatment	% of treated	% of of total
38×10^6	Primary-intermediate	37	20
33×10^6	Activated sludge	32	18
23×10^6	Trickling filter	22	12
10×10^6	Other	9	6

TREATMENT CLASSIFICATION BY FACILITIES FOR YEAR 1962

Number	Type	% of total
11,655	Total	100
2,277	Raw discharge	19.5
9,378	Treated discharge	80.5
2,794	Primary-intermediate	24.0
6,584	Secondary	56.5
800	Activated sludge (12% of secondary)	7.0
3,506	Trickling filters (53% of secondary)	30.0
1,348	Stabilization ponds (20% of secondary)	12.0

DRAINAGE BASIN WASTE DISTRIBUTION FOR YEAR 1962

Basin	% of total discharge Raw	Treated
Lake Erie	23.6	74.4
Upper Mississippi	12.3	87.7
Western Great Lakes	7.8	92.2

PHOSPHATE REPORTING

Today's investigators in sanitary engineering express the results of phosphorus analysis as P (phosphorus) and not as PO_4 (phosphate) and P_2O_5 (phosphorus pentoxide). The relationship between various methods of expression are:

$$1/mg/1 \text{ P} = 2.29 \text{ mg/1 } P_2O_5 = 3.06 \text{ mg/1 } PO_4$$
$$1 \text{ mg/1 } PO_4 = 0.75 \text{ mg/1 } P_2O_5 = 0.33 \text{ mg/1 P}$$
$$1 \text{ mg/1 } P_2O_5 = 1.34 \text{ mg/1 } PO_4 = 0.44 \text{ mg/1 P}$$

DIGESTER RECYCLE

Normal digester operation calls for recycling the supernatant from the digester into the influent line of the waste treatment plant. This procedure adds a heavy load of soluble phosphates as approximately 80% of the phosphates present in sludge are solubilized during digestion. A sizable reduction in effluent phosphate level can be realized by eliminating this addition. Wasted sludge is the only way phosphates can be removed in a purely biological process. It cannot be emphasized enough that regardless of the complex internal mechanism of phosphate removal (precipitation, adsorption, etc.) the actual phosphorus removal that a plant can achieve will ultimately depend upon the amount of sludge wasted. If digester supernatant is returned to the treatment plant, the only phosphorus removed will be that associated with the digested sludge.

PHOSPHATE REMOVAL BY WASTEWATER TREATMENT PLANTS

Type	% removal
Primary sedimentation	5-15
Extended aeration (sludge wasting)	8-15
Trickling filter	20-30
Activated sludge	30-50

ENHANCING PHOSPHATE REMOVAL

The most economical and desirable method of increasing phosphate removal would be to induce a high degree of phosphate uptake by activated sludge. The literature indicates that several factors exert an influence on phosphate removal. The most commonly considered variables are:

1. Aeration time and rate of air supply.

The rate of aeration and the aeration time have been indicated by most investigators as the most important criteria. The rate of air supply probably being the more critical of the two. Aeration rates in the order of 3 to 7 cfm/gal and detention times of 4 to 6 hours appear to be desirable.

2. Concentration of mixed liquor suspended solids.

There is some disagreement in the literature with respect to the optimum concentration of mixed liquor suspended solids (MLSS). Apparently, increased uptake has been attained at both low and high MLSS. It would appear desirable to investigate a rather broad range of solids in future studies.

3. Concentration of dissolved oxygen in aerator.

Where increased phosphate removals have been experienced, they did not occur until 1.5 to 2.0 mg/l DO had been attained. It seems essential that at least a DO level of 2 mg/l be maintained in the last half of the aeration tank to insure that solids retain phosphate enrichment through secondary clarifiers.

4. Time of sludge retention in secondary clarifier.

Phosphate uptake by sludge organisms in the absence of growth leaks out when the dissolved oxygen level falls. Leakage will occur in

the secondary settling tank when the sludge consumes available dissolved oxygen. It has been suggested that solids detention time in final clarifiers should be less than 30 minutes. If rapid sludge withdrawal is important, we should design a more efficient and rapid method of separating solids from liquids.

SEMINAR ON PHOSPHORUS REMOVAL
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PHOSPHORUS REMOVAL BY TERTIARY TREATMENT
WITH LIME AND ALUM

C. A. Brunner

I. Lime Treatment of Secondary Effluent

A. Removal is probably by formation of insoluble compounds containing calcium and phosphate such as hydroxyapatite.

B. Equipment is the same as that used for lime softening of water. Upflow clarifiers with recycle of sludge to the flocculating zone work well. Single-stage equipment, i.e., equipment allowing operation at only one pH, gives good phosphate removal. Two-stage equipment, i.e., equipment allowing operation at one pH followed by operation at a lower pH is presently being tested. Phosphate removal should be excellent. A two-stage system is inherently more complicated than the single-stage system and may have significantly higher capital cost.

Filtration following settling improves phosphate removal. Good removal can be obtained, however, without the use of filters.

C. Lime dose for phosphate removal cannot be predicted without a knowledge of the concentrations of other materials in the water, especially bicarbonate alkalinity. If the phosphate is removed as hydroxyapatite, only about 3.0 parts by weight of CaO are required per part of phosphorus. That amount of lime constitutes only a minor fraction of the total usually required for the numerous reactions that occur during lime treatment.

Experimental results indicate phosphate removal improves with increasing pH. For single-stage treatment, a clarifier pH of about 9.5 has been found to give 80-percent removal of phosphate in one pilot study with high alkalinity water. The average lime dose for that study is estimated at 250 mg/l as $Ca(OH)_2$.

D. The pH of lime treated water is likely to be too high for direct discharge. Recarbonation will be necessary in these cases.

E. The sludge formed during treatment of a high alkalinity water is granular and of high density. It consists largely of calcium carbonate. Solids concentrations in the sludge of more than 10 percent can be obtained. Settling is rapid and can be improved by recycle of part of the sludge to the flocculating zone of the clarifier. Overflow rates of 1400 gpd/sq ft have been shown to be satisfactory for good phosphate removal without using filtration.

The sludge from waters of low mineral content is more flocculent and may not settle well especially if there is appreciable suspended material left in the wastewater after biological treatment. Experimental work with this type of water has shown that addition of sodium carbonate or some other form of alkalinity may be necessary to increase floc density. Iron salts also improve operation. Settler overflow rates will be lower than for the highly mineralized waters. A study is being made to determine methods for improving lime treatment of low-mineral-content waters.

F. For large plants treating high alkalinity water, calcination appears most appropriate for sludge disposal. Usable lime is recovered in the process.

II. Alum Treatment of Secondary Effluent

A. Phosphate removal is either by incorporation in a complex along with aluminum and hydroxyl groups or by adsorption on aluminum hydroxide floc.

B. Equipment should be the same as that used for alum clarification. Pilot work has shown a horizontal flocculator-settler arrangement to be satisfactory. Use of filters after the settler will give improved phosphate removal, but these should not be necessary to achieve 80-percent removal.

C. Alum dose is difficult to predict at present. The minimum that appears necessary is two parts of aluminum by weight per part of phosphorus. For very high degrees of phosphate removal the ratio is four or more. For municipal secondary effluent the required dose of commercial alum is likely to be 200 mg/l or more for 80-percent phosphate removal.

D. The pH of most secondary effluent is in a reasonable range for alum treatment without adjustment. The addition of alum will lower the pH. If the water contains enough alkalinity, the change will be minor. If there is not enough alkalinity to buffer the water, raising of the pH may be desirable before discharge. One way to overcome large changes in pH is to supply part of the alumina in the form of sodium aluminate. Experimental work has shown a combination of alum and sodium aluminate to give results comparable with alum alone.

E. The sludge formed is voluminous and of low density. As a result, settler overflow rates as low as 700 gpd/sq ft may be necessary in the absence of filtration to prevent excessive floc carry-over in the treated water. Addition of other chemicals such as activated silica may be required to improve settling. Since large volumes of sludge are produced sludge handling presents problems.

F. The economics of alum treatment would be improved if the aluminum could be recovered and reused. Both chemical costs and sludge disposal costs would be decreased. Although several schemes have been investigated for aluminum recovery, none have yet been found to be practical.

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RECOVERY AND DISPOSAL OF ALUM SLUDGES

by

J. B. Farrell

Presented at the

Conference on Removal of Phosphate
Chicago, Illinois
May 1-2, 1968

EFFECT OF pH ON PHOSPHATE REMOVAL BY ALUM

SYNTHETIC EFFLUENT

300 mg/l HCO_3^-

4–12 mg/l P

SYMBOL	Al ADDED / P INITIAL
○	0.65
□	1.09
△	1.64
◇	2.61

pH OF SUPERNATANT AFTER TREATMENT

FRACTION OF P REMOVED $\frac{Pi-Pt}{Pi}$
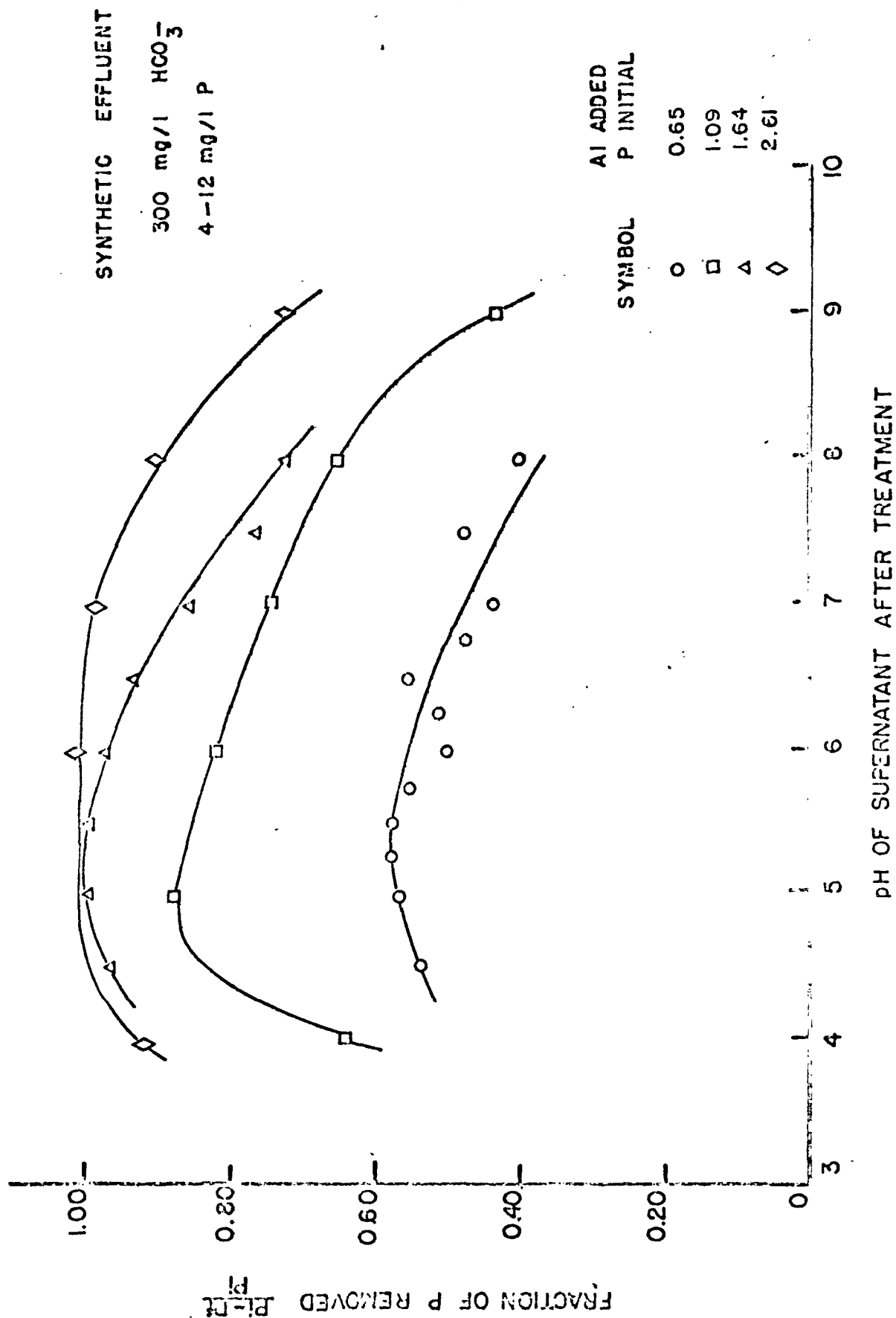
DISPOSAL OR UTILIZATION

OF ALUM-PHOSPHATE SLUDGES

1) LANDFILL

2) USE AS A CHEMICAL RAW MATERIAL

   a) LUBRICANT IN GRINDING MILLS FOR FIRECLAY MANUFACTURE

   b) RUBBER FILLER

   c) SEPARATION OF PHOSPHATE FOR FERTILIZER USE

3) RE-SOLUTION OF AL AND REUSE IN TERTIARY TREATMENT

   a) ACID PROCESS

   b) BASIC PROCESS

RECOVERY PROCESSES

I. ACID PROCESSES

   a) DISSOLVE SLUDGE IN H_2SO_4

      (1) REMOVE PO_4 BY ANION EXCHANGE

      (2) REMOVE Al BY CATION EXCHANGE

      (3) REMOVE PO_4 BY CHEMICAL OR PHYSICAL MEANS

II. BASIC PROCESSES

   a) DISSOLVE SLUDGE IN NaOH, PRECIPITATE PO_4 WITH $CaCl_2$

   b) DISSOLVE SLUDGE IN $Ca(OH)_2$ AT ROOM TEMP., HIGH TEMP.

RECOVERY OF ALUM BY AN ACID-FREEZING PROCESS:
DAER ALUM RECOVERY PLANT

SLUDGE FROM SEDIMENTATION
TANKS AND FILTERS

100,000 gpd
(0.5 % SOLIDS)

HOLD TANK

ACID TREATMENT

RECOVERED ALUM

H2SO4

SLUDGE

THICKENER

SLUDGE

REFRIGERATION PLANT

THAWED SLUDGE
(1200 gpd)
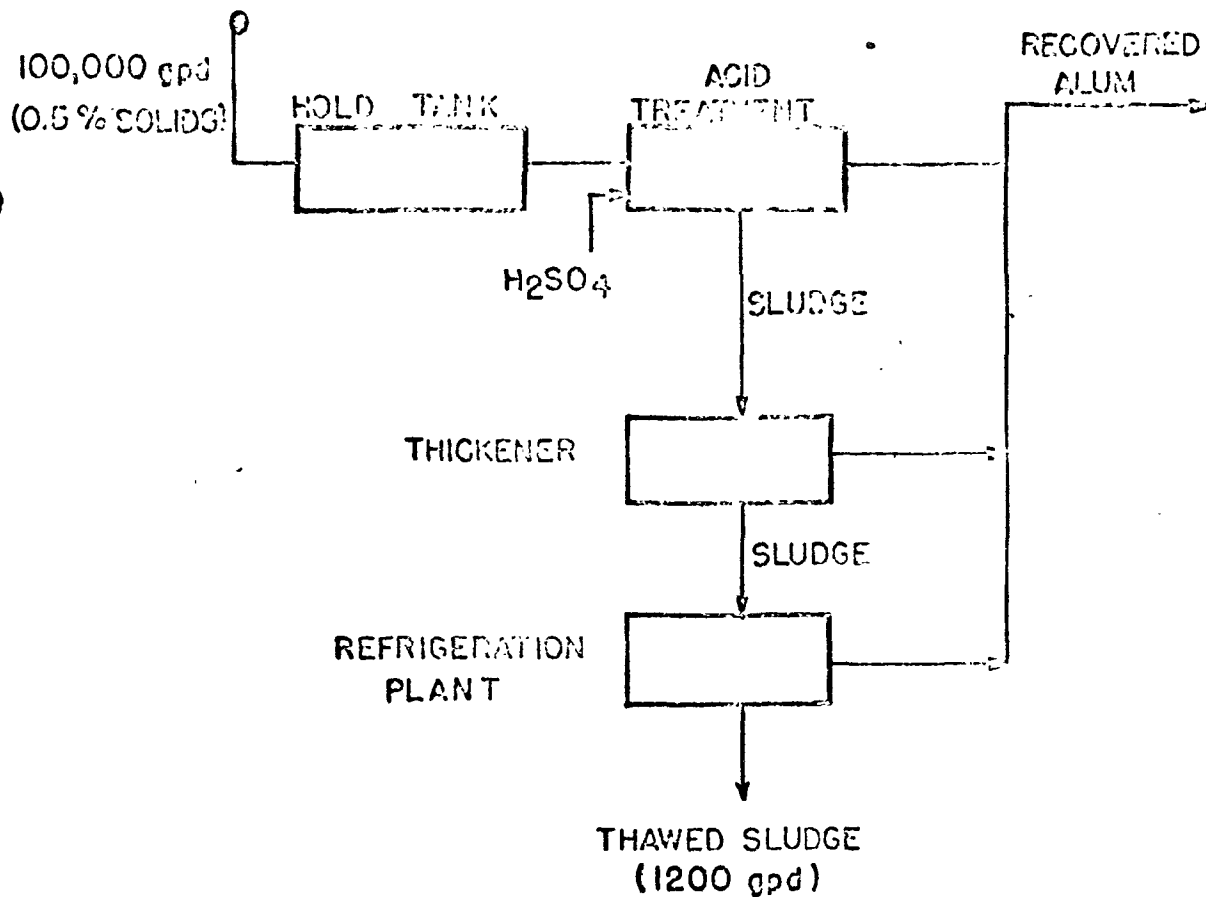
PHOSPHATE REMOVAL BY ALUM PRECIPITATION, WITH AL RECOVERY BY NaOH – CaCl₂ PROCESS

SECONDARY EFFLUENT

H_2SO_4

MAKEUP ALUM

MIXER

FLOCCULATOR

CLARIFIER

TO WATERCOURSE

RECOVERED ALUMINATE

$CaCl_2$

CLARIFIER   MIXER   REACTOR   MIXER

NaOH

$Ca_5(OH)(PO_4)_3$ SLUDGE AND UNRECOVERED Al
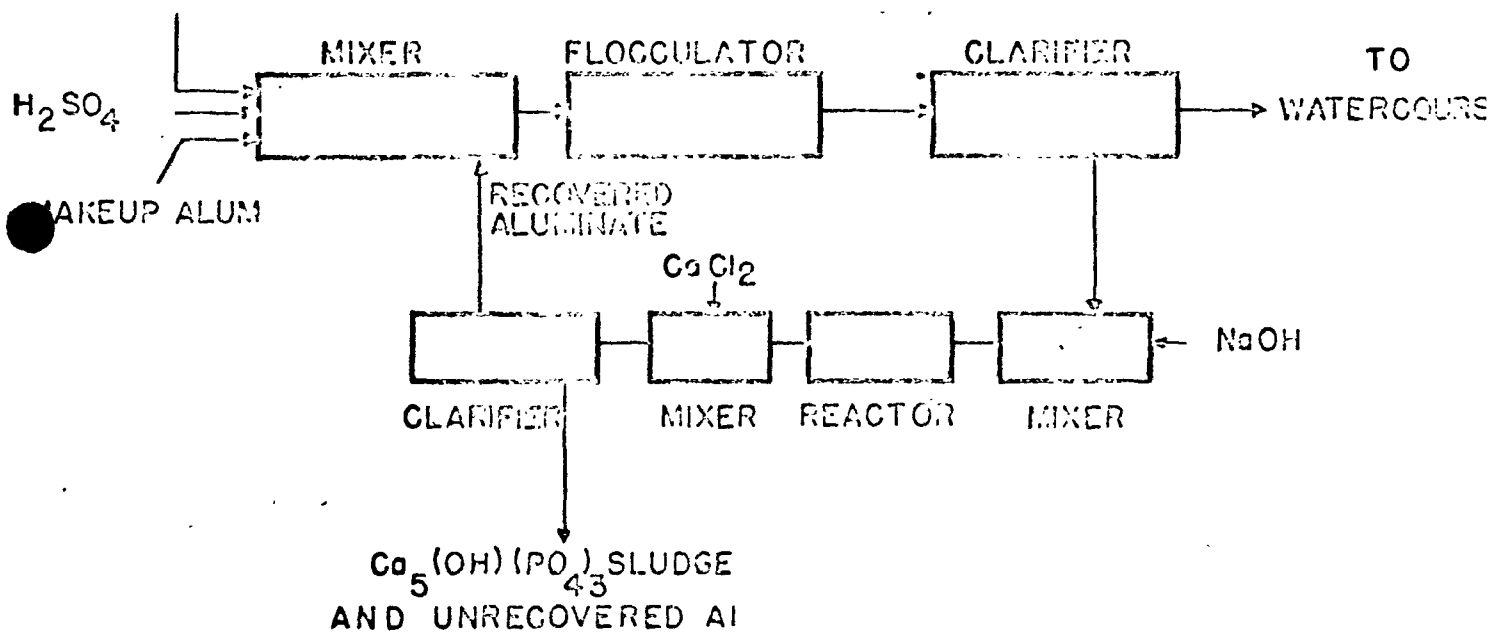
COMPARISON OF CHEMICAL COSTS FOR RECOVERY

OF AL FROM ALUM-PHOSPHATE SLUDGES

COST OF CHEMICALS, ¢/1000 GAL*

	ALUM	NaOH	CaCl$_2$	Ca(OH)$_2$	TOTAL
NaOH-CaCl$_2$					
R-L-K	1.18	0.81	0.50		2.49
CULP & SLECHTA	0.97	2.13	1.38		4.48
LIME ONLY					
50% RECOVERY	2.0			0.62	2.62
NO RECOVERY	4.0				4.0

* 167 mg/l AL DOSE (AS ALUM, MOL.WT. OF 600)

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CHEMICAL-PHYSICAL METHODS FOR CONTROL

AND REMOVAL OF NITROGEN

by

J. B. Farrell

CHEMICAL AND PHYSICAL METHODS

FOR NITROGEN REMOVAL

1) PHYSICAL

   a) ADSORPTION

   b) COAGULATION

   c) STRIPPING

2) CHEMICAL

   a) ION EXCHANGE

   b) OXIDATION

COMPARISON OF SOLUBILITY OF
CARBON DIOXIDE AND NH_3 IN WATER

HENRY'S LAW CONSTANT (H)

TEMP. (°C)	CO_2	NH_3
20	1,420	0.7
40	2,330	1.8
60	3,410	3.8

WHERE $p_a = Hx_a$

p_a = VAPOR PRESSURE OF DISSOLVED SUBSTANCE "a" (atm)

x = MOLE FRACTION OF "a"

H = HENRY'S LAW CONSTANT

EFFECT OF ION EXCHANGE RESIN SELECTIVITY ON REMOVAL OF A GIVEN ION

PROBLEM: REMOVE B FROM SOLUTION CONTAINING A, B, AND C.
RESIN INITIALLY IS IN FORM C·R.

CASE 1	CASE 2	CASE 3
A > B > C	B > A > C	A > C > B
A	B	A
B	A	B,C
C	C	
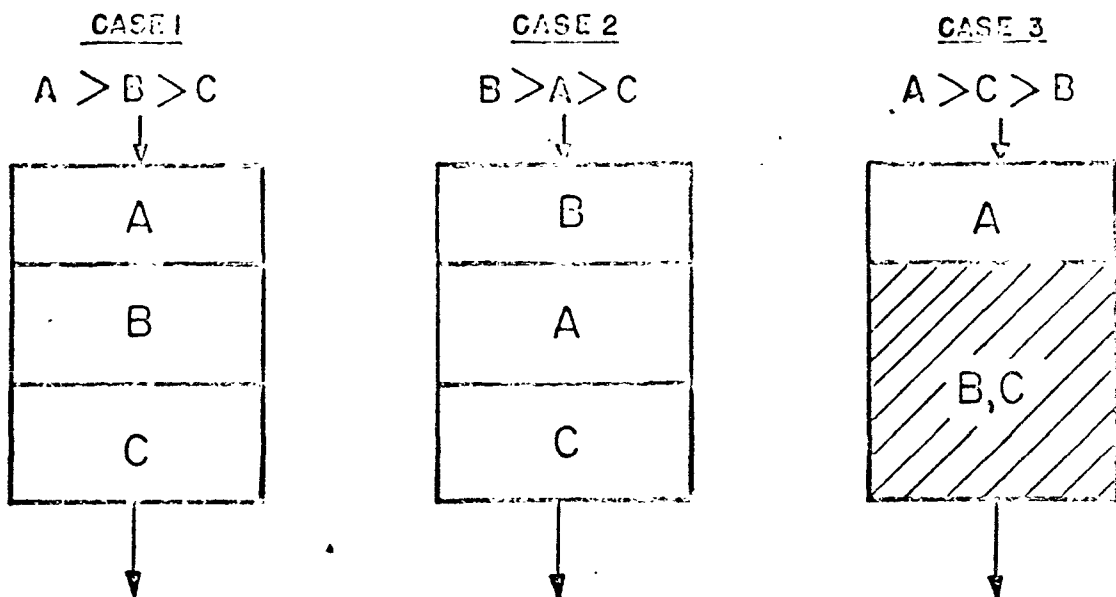
APPROXIMATE COST OF REMOVING

90 PERCENT OF NITROGEN[1]

	¢/1000 GAL.
AMMONIA STRIPPING	2.3 (4.9)
ION EXCHANGE	
a). REMOVAL OF NH_4 [2]	8.1
b) REMOVAL OF NO_3	18.0
CHLORINATION[2]	5.3

(1) PLANT SCALE IS 10 MGD

(2) 20 mg/l N IN FEED

LIST OF PERTINENT REFERENCES

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MANAGING CONTINUOUS FLOW BIOLOGICAL DENITRIFICATION

E. F. Barth and M. B. Ettinger, Research Chemists,
Federal Water Pollution Control Administration,
Cincinnati Research Laboratory, Cincinnati, Ohio

This report is a summary of our current research aimed at modification of wastewater treatment processes to efficiently remove nitrogen via a biological process.

Figure 1 is a simple flow sheet of the nitrogen cycle during wastewater treatment. Several points are evident. First, if conditions are not proper for nitrification to occur, any ammonia in excess of that needed for cellular synthesis will be discharged in the final effluent, and very little nitrogen removal will be observed. Second, for denitrification to occur nitrification must be controlled, and a source of organic carbon must be available. The reaction of nitrite with a primary amine is the classical Van Slyke reaction and does not occur to any great extent in biological systems. This figure illustrates the importance of understanding the sequence of events when attempting to manage a complex biological system.

Our early studies to control denitrification were attempts to modify existing structures to accomplish efficient nitrogen removal. Figure 2 is the first modification tried. The air diffusers from the first two sections of a conventionally operated 100-gallon-a-day activated sludge pilot plant were removed, and the sludge kept in suspension by stirring. In all the studies reported here the sludges produced during treatment were digested anaerobically and the supernatant returned to the process at the primary settler. All nitrogen removals, unless otherwise stated, are overall removals. A pump to recycle mixed liquor (100 percent of influent flow) was installed between the aerator exit and the final settler. The rationale of this system was to recycle the active solids from the mixed liquor with the associated nitrates to the anaerobic section where the organic carbon in the primary effluent would force denitrification to occur. Nitrogen removal was encouraging but varied widely; however, with this modification nitrate would always be discharged in the effluent because of the split stream at the mixed liquor exit. We therefore tried systems that would treat the entire secondary effluent flow.

Figure 3 is a system of alternate aerobic and anaerobic sections. Because of the short detention time in each section and the cycling of the organisms between aerobic and anaerobic conditions, process control was unstable and the ability to nitrify was lost. The only nitrogen removal was with the primary sludge.

Figure 4 illustrates the next attempt to accomplish controlled denitrification. A holding tank was installed between the mixed liquor exit and the final settler to provide detention time for denitrification to occur. As indicated, nitrogen removals were not much better than the original system in Figure 2. Only about 50 percent

of the oxidized nitrogen was denitrified in a six-hour period. This was because at the mixed liquor exit the organisms had a very low respiration rate, and the major fraction of the organic carbon had been degraded during the conventional aeration ahead of the denitrification tank. We tried various additives to this tank to serve as a source of organic carbon, such as primary effluent, primary sludge, and glucose. Each proved to have disadvantages as listed in Table I.

After considering these results, it was apparent that to control nitrification, denitrification, and keep detention times as short as possible a system such as shown in Figure 5 would be necessary. We are currently working with this system in a 200-gallon-per-day pilot plant. There are three separate sludge systems, each operated as a separate but interrelated unit process.

The high rate sludge system removes about 80 percent of the influent carbon as cellular material and carbon dioxide. The effluent from this unit containing residual carbon and the bulk of the influent nitrogen, in the form of ammonia, goes to a nitrification system. Since the majority of the carbon has been removed, danger of washout of nitrifiers by excessive sludge washing is eliminated. This unit has its own sludge system and, therefore, can be considered an enriched culture of nitrifiers. Nitrification is accomplished in three hours. The high quality nitrified effluent then goes to the denitrification section. We have found it necessary to add an organic carbon source at this point. Methyl alcohol has been tried and found to be very satisfactory. The organisms oxidize rather than synthesize this material, and residual methyl alcohol has not been found in the effluent. As shown in the last entry of Table I denitrification efficiency is high. Overall nitrogen removals of 85 percent have been obtained. However, the process is not under complete control, as indicated by the spread in the removal efficiencies. Much of the trouble in process control is due to the pilot scale size of the operation; management of a three-sludge system with small sludge lines and limited linear flows is difficult. It is planned to study this system in a 0.5 mgd demonstration plant.
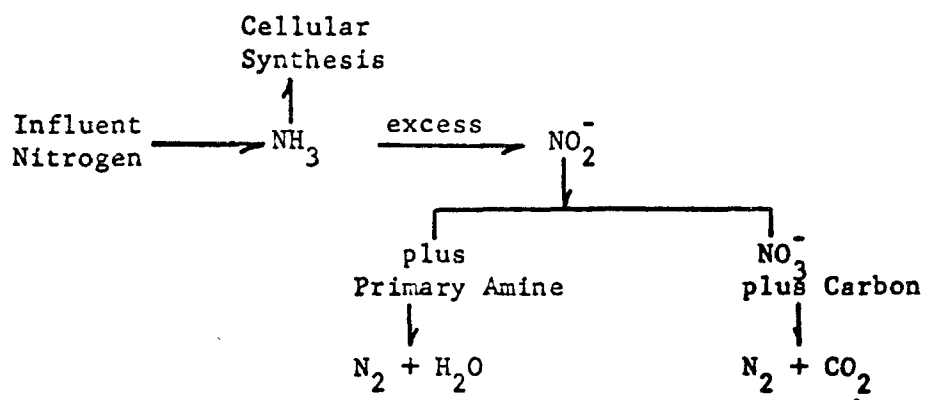
FIGURE 1

PERCENT
NITROGEN
REMOVAL

30-70%

0-20%

30-60%

FIGURE 2

STIR
NO AIR   AIR

PRIMARY

MIXED LIQUOR RECYCLE
RETURN SLUDGE

FIGURE 3

AIR   NO AIR   AIR   NO AIR   AIR   NO AIR

PRIMARY

RETURN SLUDGE

FIGURE 4

ANAEROBIC
STIR

AEROBIC

PRIMARY

RETURN SLUDGE

FIGURE 5

PRIMARY

HIGH RATE
C→CO₂

RETURN SLUDGE

NITRIFICATION
NH₃→NO₃

RETURN SLUDGE

DENITRIFICATION
NO₃→N₂

RETURN SLUDGE
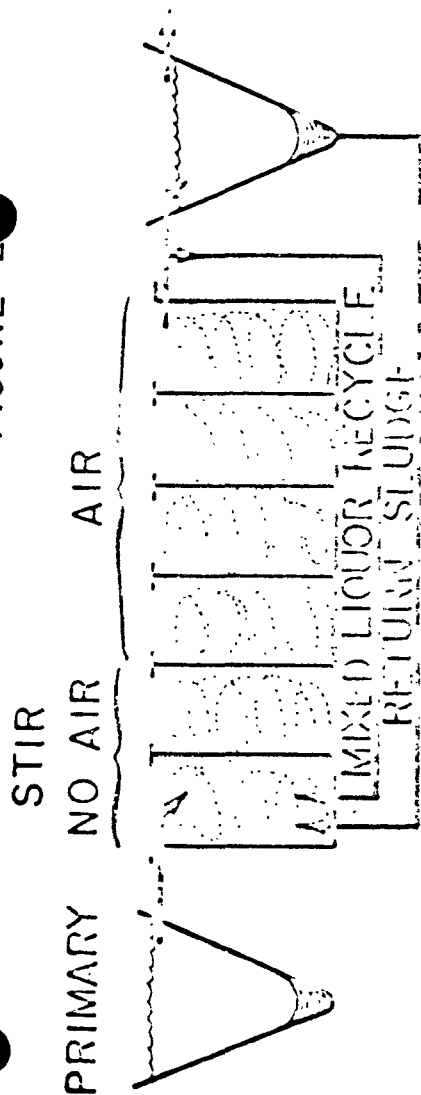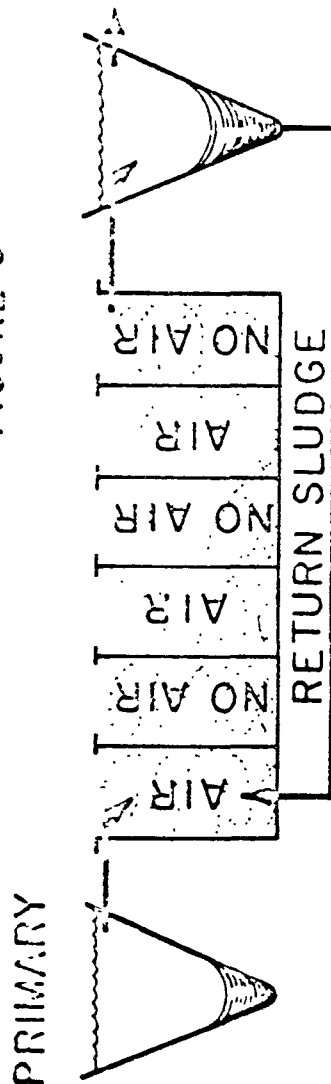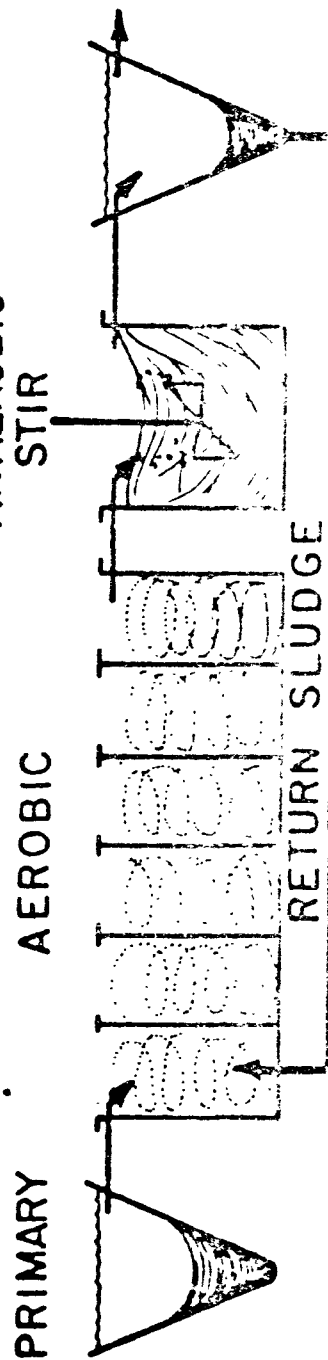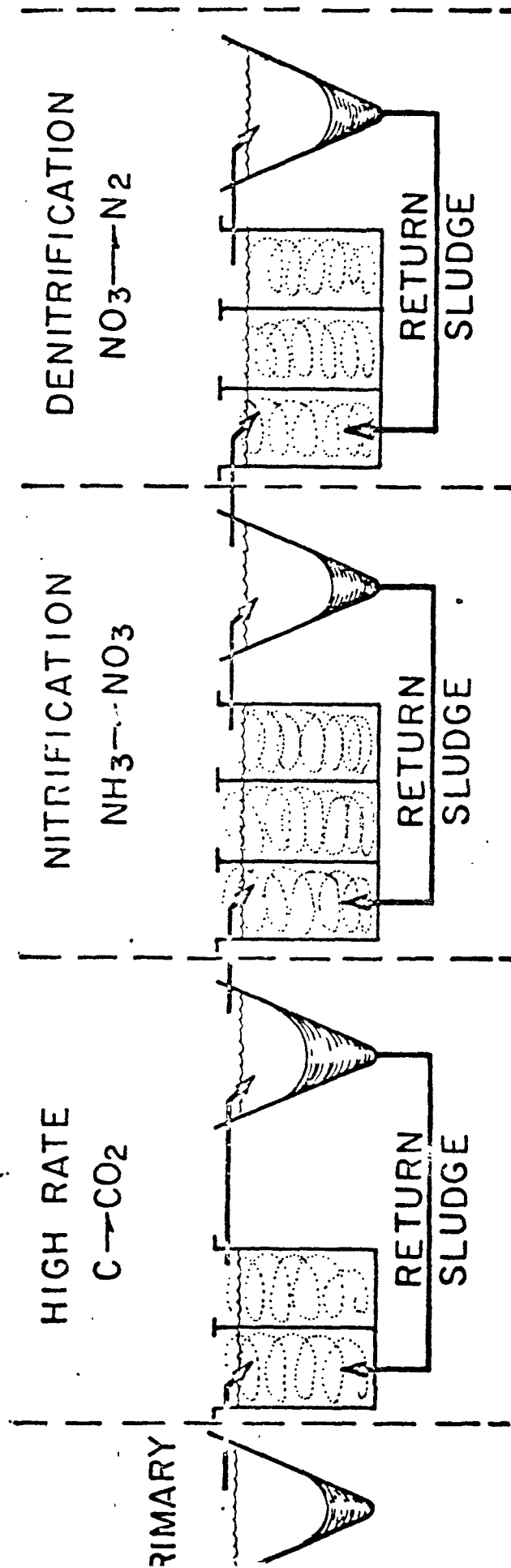
TABLE I

Denitrification Scheme	Reason	Percent Removal of Oxidized N	Disadvantages
Nitrified effluent plus high rate sludge	Sorbed organics Active sludge	50-90%	TKN (solids), NH_3, and turbidity
Nitrifying sludge plus nitrified effluent	Simple flow sheet, no bleed through of nitrogen	10-30%	Six hour detention time. Sludge inactive, no organic to force denitrification.
Primary effluent to denitrification stage	Source of organics Simple flow sheet	50-60%	NH_3 and turbidity
Primary sludge to denitrification stage	Reduce hydraulic flow, source of organics	50-60%	TKN (solids), NH_3, and turbidity
Glucose fed to denitrification stage	Source of organic No bleed through of nitrogen	20%	Synthesis instead of oxidation light bulking sludge
Methanol fed to denitrification stage Three separate sludge systems	Cheap organic oxidation instead of synthesis	75-90%	Slight turbidity Process control

ALTERNATIVE METHODS OF PHOSPHORUS REMOVAL[*]

Jesse M. Cohen

Introduction

1. At the present time, reagents based on iron, aluminum or lime are the chemicals of choice for removal of phosphates from wastewater.

2. Eventually, a variety of reagents and methods will be available to provide a wide choice of design and performance for removal of phosphates.

3. None of the methods discussed here are now ready for practical application.

Activated Alumina

1. Activated alumina is a synthetic material consisting largely of aluminum oxide and available in a variety of forms ranging from powders to granules.

2. The unique properties are its high specific surface area, 200-400 m^2/gram, and its ability to selectively sorb phosphates. Moreover, when operated in a column mode, alumina can be loaded and regenerated for a great number of cycles with only 5-8% loss of alumina per cycle.

3. Virtually complete, > 99%, removal of phosphates is obtained. A feed phosphate concentration of 25 mg/l showed no detectable phosphate in the product until more than 1000 bed volumes have passed.

4. Capacity to remove polyphosphates is better than for orthophosphate.

5. As a possible polishing process, better than 99% removal can be obtained from a 1 mg/l solution while producing 20,000 bed volumes of product.

6. Special advantages are: No dissolved solids are added to product, no pH changes occur, feedwater composition and quality have little effect on removal efficiency, residual phosphates are very low, $\approx$ 0.05 mg/l of phosphate.

7. Chemical costs are estimated as 3.9 cents/1000 gals for a feed concentration of 14 mg/l of phosphate.

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[*]Talk delivered to Workshop on Phosphorus Removal, Chicago, Illinois, May 1 & 2, 1968.

Lanthanum

1. Based on solubility of its phosphate salt, lanthanum must be considered as a possible precipitant.

2. Theoretical solubilities are not obtained with iron or aluminum because of competing hydrolysis reactions which reduce the metal ions available for phosphate precipitation.

3. Studies have shown that lanthanum is only slightly hydrolyzed and shows less tendency to form soluble complexes with phosphate.

4. Desk top considerations predict that lanthanum can be recovered by treatment with alkali and then reused.

5. A recent price quotation of 60¢ per pound of lanthanum in a lanthanum-rich mixture of rare earths (down from $3.00/lb) is encouraging.

Ion Exchange

1. Ion exchange materials, both natural and synthetic have potential for phosphate removal.

2. An important group of naturally occurring ion-exchange materials are the zeolites and clay minerals.

3. Anion exchange capacities of clays are too small to be economically useful, but chemical modifications can produce increased capacities.

4. Process based on chemically modified clay minerals can become economically attractive. Clays are abundantly available and cheap - just cents per pound.

5. Synthetic ion-exchange materials have adequate anion exchange capacities but poor selectivity.

6. Since resins exchange both sulfate and phosphate about equally well, capacity must be shared with these anions. The problem is that wastewater contains 2-5 times as much sulfate as phosphate.

7. A synthetic ion-exchange resin with high selectivity for phosphate anion is needed.

Soil Systems

1. Soil systems, as treatment devices rather than as means of disposal, are potentially cheap methods of removing phosphates.

2. Soil treatment mechanisms include biological oxidation, adsorption, chemical oxidation, chemical precipitation, ion exchange and plant assimilation.

3. Potential of soil to remove phosphate is enormous. During 6 years of field study, 1.6 tons of phosphorus was fixed in the upper 6 inches of an acre of soil. Phosphate capacities of some soils range from 11.2 to 40 tons to as high as 205 tons of phosphorus per acre half-foot. Operating a soil system at 1 gal/d/sq ft, phosphorus capacity would not be exhausted for upwards of a century.

4. Phosphate fixation in soil can be attributed to many factors such as anion exchange, adsorption, chemical precipitation and plant assimilation.

Reverse Osmosis

1. Reverse osmosis is a process which uses pressure and semipermeable membranes to force water to be transported through the membrane leaving salts and other molecules behind.

2. This process is being developed principally as a device to remove contaminants in general.

3. Membranes show selectivity for ions proportional to valence; hence, multivalent phosphate and sulfate are more completely removed than chloride.

4. If reverse osmosis can be developed to an economically useful process for partial demineralization, it is useful to know that phosphate will be almost completely removed.

Up-flow Clarification

1. Sludge-blanket or up-flow clarification offers an alternative device to horizontal-flow flocculators and sedimentation basins.

2. Principal advantages are: reduced capital costs, reduced land requirements, reduced detention time, multi-functional processes in a single unit.

3.   Ability to remove pollutants such as phosphates, suspended solids, color and organic solids, is equal to or greater than horizontal equipment. Intimate contact with sludge blanket enhances chemical and physical reactions.

4.   Preliminary laboratory results show 80-97% removal of phosphates with blanket depths of 4-8 feet using alum, ferric sulfate or lime.

5.   Rates of 10-15 gal/hr/sq ft which provide detention times of about 1 hour may be used.

Concluding Remarks

1.   Alternative methods for removal of phosphate are being considered.

2.   None of these alternatives are now ready for application.

3.   Research will almost certainly make some of these alternatives useful for application in specialized instances.

COST OF REMOVING PHOSPHORUS FROM WASTEWATER

By Robert Smith
Seminar on Phosphorus Removal
Chicago, Illinois - May 1 & 2, 1968

It has been shown [1-6] that phosphorus can be effectively removed from wastewater by adding chemicals such as lime, alum, ferric sulfate, or sodium aluminate at various points within the conventional wastewater treatment plant or in a separate coagulation and sedimentation step downstream of the conventional plant. The two principal cost items associated with phosphorus removal are the cost of chemicals and the cost of disposal of the waste sludge produced.

Reasonably reliable information is available for estimating the cost of phosphorus removal at a specific site when the characteristics of the wastewater, the cost of delivered chemicals, and the requirements for sludge disposal are known. Cost estimates for hypothetical plants are also useful to indicate, roughly, limits of total cost that might be encountered as a result of installing and operating processes for phosphorus removal. Estimated costs for specific items associated with phosphorus removal together with some guarded estimates for hypothetical plants are given below.

ADDITION OF ALUM OR SODIUM ALUMINATE TO THE AERATOR

Barth and Ettinger[1] have shown that the concentration of phosphorus in the effluent stream from the activated sludge process can be reduced from 10 mg/l to about 0.5 mg/l by adding one part of aluminum to the aerator influent stream for each part of phosphorus present in the aerator influent stream. The aluminum can be added to the aerator in the form of aluminum sulfate (alum) or sodium aluminate both of which are available in liquid or dry form. In the range of plants between 1 mgd and 100 mgd the cost of purchasing the necessary feeding equipment and storage tanks amounts to only 0.01 - 0.10 cents per 1000 gallons of water treated. Operating and maintenance cost has not been estimated. The cost of chemicals delivered to a 100 mgd plant in Columbus, Ohio has been estimated to range between 2.0 and 2.5 cents/1000 gallons for alum and 3.0 - 3.5 cents/1000 gallons for aluminate. Barth and Ettinger[1] have also shown that the phosphorus removed with the waste activated sludge is retained by the sludge even after the sludge has been subjected to anaerobic decomposition for periods far in excess to the normal detention times. Thus, no abnormal amount of phosphorus will be returned to the activated sludge process with the digester supernatant. Barth and Ettinger[1] found that the addition of aluminum to the aerator has a beneficial effect on the settling characteristics of the sludge but no marked improvement of the performance of the activated sludge process in removing organics was noted.

SEPARATE LIME CLARIFICATION PROCESS

Coagulation and sedimentation following lime addition in a separate process downstream of the activated sludge process is somewhat more expensive than adding chemicals to the aerator. The lime clarification process, however, removes a significant portion of the remaining suspended solids and BOD along with the phosphorus. Another advantage of the lime clarification process is that ammonia

ripping, which requires raising the pH of the water, can be used downstream of the process to remove ammonia from the water.

A high-density solids-contact process similar to the Infilco Densator* is believed to be the most economical lime clarification process. Waste lime sludge from the Densator has a density of about 300 grams/liter. A separate sludge thickening step is, therefore, not required.

The installed cost of Densators and related equipment was supplied[7] by Infilco/GATX for plants in the range of 250-450 mgd and for the range of 0.07 - 0.50 mgd. These cost estimates which are based on using an overflow rate of 2000 gpd/sq. ft. are shown in Figure 1.

Land required for a Densator installation is about 40 acres for the 450 mgd size. Taking the cost of land to be $5000 per acre this represents about one tenth of a cent per 1000 gallons of water treated.

Debt Service charges (4-1/2% - 25 yr.) expressed as cents per 1000 gallons of water treated are shown in Table I.

The cost of operating and maintenance labor was also estimated by Infilco personnel as about $260 per year per mgd of flow at the 450 mgd size. This is based on four men on each of three shifts at a cost of $9800 per man-year. This estimate for operating and maintenance cost was extrapolated to smaller sized plants by means of a log-log plot having a slope of (-1/3). At the very small plant sizes the cost estimated in this way might be low.

The cost of operation and maintenance for the Densator installation expressed in cents per 1000 gallons of water treated is shown in Table I.

The lime dose required will depend on the alkalinity of the water and the target pH to be achieved or the fraction of phosphorus to be removed. Since no specific site is being considered a target pH of 11.0 and a lime dose of 227 mg/l of quick-lime (CaO) or 300 mg/l of hydrated lime will be used for estimating lime cost. The cost of purchasing lime in the Chicago, Ill. area is about $18.50 per ton when delivered in lots of 50 tons or more. The amount of quicklime required by the process is about one ton per day per mgd. The cost of purchasing lime, therefore, amounts to about 1.75 cents per 1000 gallons of water treated. Infilco also recommends the use of 50 mg/l of ferrous sulfate to help coagulate and settle the fines. The cost of ferrous sulfate delivered is about $41.75 per ton or about 0.87 cents per 1000 gallons of water treated.

Disposal of the waste lime sludge presents a problem which is greatly affected by the disposal means available, the cost of recalcination, and the amount of waste lime sludge produced. For example, if the plant is small and disposal sites are close and available the most economical method for disposal might be trucking to a landfill.

* Mention of proprietary equipment does not constitute endorsement by FWPCA.

On the other hand, when the amount of sludge produced is large and the disposal sites are at a great distance, recalcination and reuse of the lime is likely to be the most economical solution.

In Cincinnati the cost of hauling sludge by truck was found to be 32 cents per ton-mile for a one mile one-way trip and 5 cents per ton-mile or less for one-way trips of more than 24 miles. For a 25 mile one-way trip the cost of sludge disposal is about 0.67 cents per 1000 gallons of water treated. Barging to sea from the Washington, D.C. area was found to cost 1.6 cents per ton-mile for a 450 mile one-way trip. This represents about 1.9 cents per 1000 gallons of water treated. A pipeline cost study[8] was made by Rand Development Corp. based on a pipe 160 miles long. If the waste sludge pumped through the pipeline is assumed to have a density of 35 grams/liter the cost of sludge disposal would be about 2.1 cents per ton-mile. Crow[9] reported a disposal to landfill cost in Gainesville, Flordia of $23 per ton of feed lime.

Another factor in computing the advantage of recalcination and reuse of lime is the hardness of the water treated and the target pH used. For example, Crow[9] reported that 1.24 lb. of quicklime had been recovered by recalcination for each pound of quicklime used in the process. When the concentration of calcium in the feedwater is low (soft water) the recovery by recalcination might be only 70% of the lime used in the process.

As the pH of the water is raised above about 10.0 increasing amounts of calcium escape in the effluent. This is shown for a specific case in Figure 2. Recovery of lime by recalcination is thus hampered.

Estimates for the cost of recalcining waste sludge from water treating plants range from $12 per ton for a 150 ton/day plant to $34 per ton for a 6 ton/day plant. The total cost of recalcination can be estimated roughly from the following relationship.

$$\text{Total Cost of Recalcination, \$/ton} = \$62/(\text{tons/day})^{.32}$$

As mentioned previously, when the design guidelines are known it is possible to make a fairly accurate preliminary cost estimate for the lime clarification process. For hypothetical plants the most reasonable approach is to assume that lime can be produced by an on-site recalcination plant at about the same cost as the cost of purchased lime. If makeup lime is required to prevent excessive build up of phosphorus and other contaminants the cost is thus unaffected. Under this assumption we can then say that the net saving accomplished by recalcining the waste lime sludge is equal to the cost of disposal by transportation to a landfill.

Total estimated cost for the lime clarification process in cents per 1000 gallons of water treated are shown in Table I and Figure 3.

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TABLE I

TOTAL COST OF PHOSPHATE REMOVAL (Cents per 1000 gallons)				
	Size of Plant			
	1 mgd	10 mgd	100 mgd	250 mgd
Capital amortization	.97	.79	.65	.59
Land amortization	.09	.09	.09	.09
Operating and maintenance	.41	.14	.08	.06
Cost of chemicals				
Lime	1.75	1.75	1.75	1.75
Iron salt	.87	.87	.87	.87
Cost of sludge disposal by hauling (to land fill (25-mile one-way trip)	.67	.67	.67	.67
TOTAL	4.76	4.31	4.11	4.03
Savings if sludge can be recalcined	-.67	-.67	-.67	-.67
TOTAL (with recalcining)	4.09	3.64	3.44	3.36

Source: FWPCA, Lake Michigan Water Pollution Enforcement Conference

CAPITAL COST FOR INFILCO DENSATORS

DOLLARS versus AVERAGE FLOW

Capital Cost, thousands of dollars
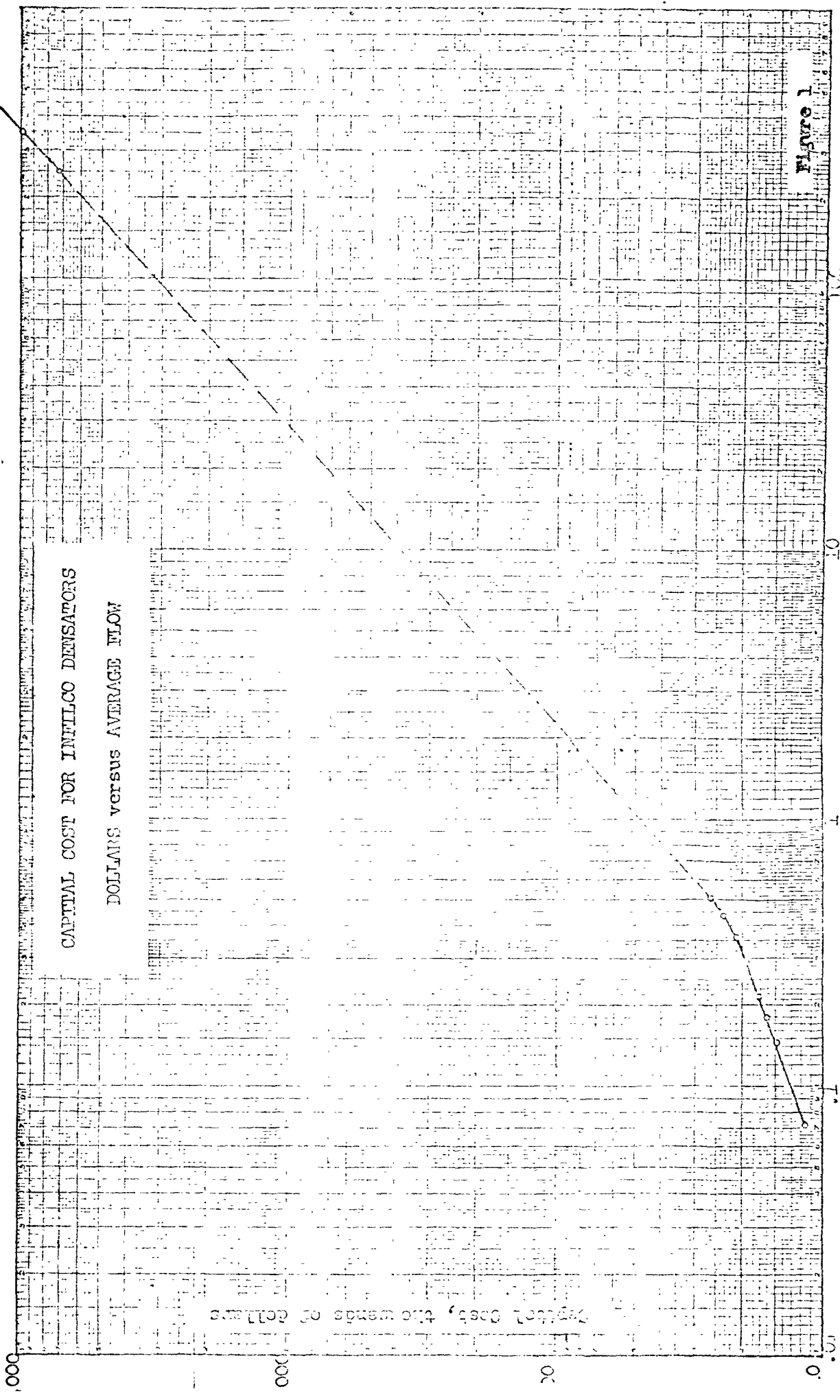
Plant Size, millions of gallons per day

Figure 1

R. Smith 4/4/67

CALCULATED QUICKLIME DOSE TO REACH ANY pH

and

CALCIUM ION IN SOLUTION
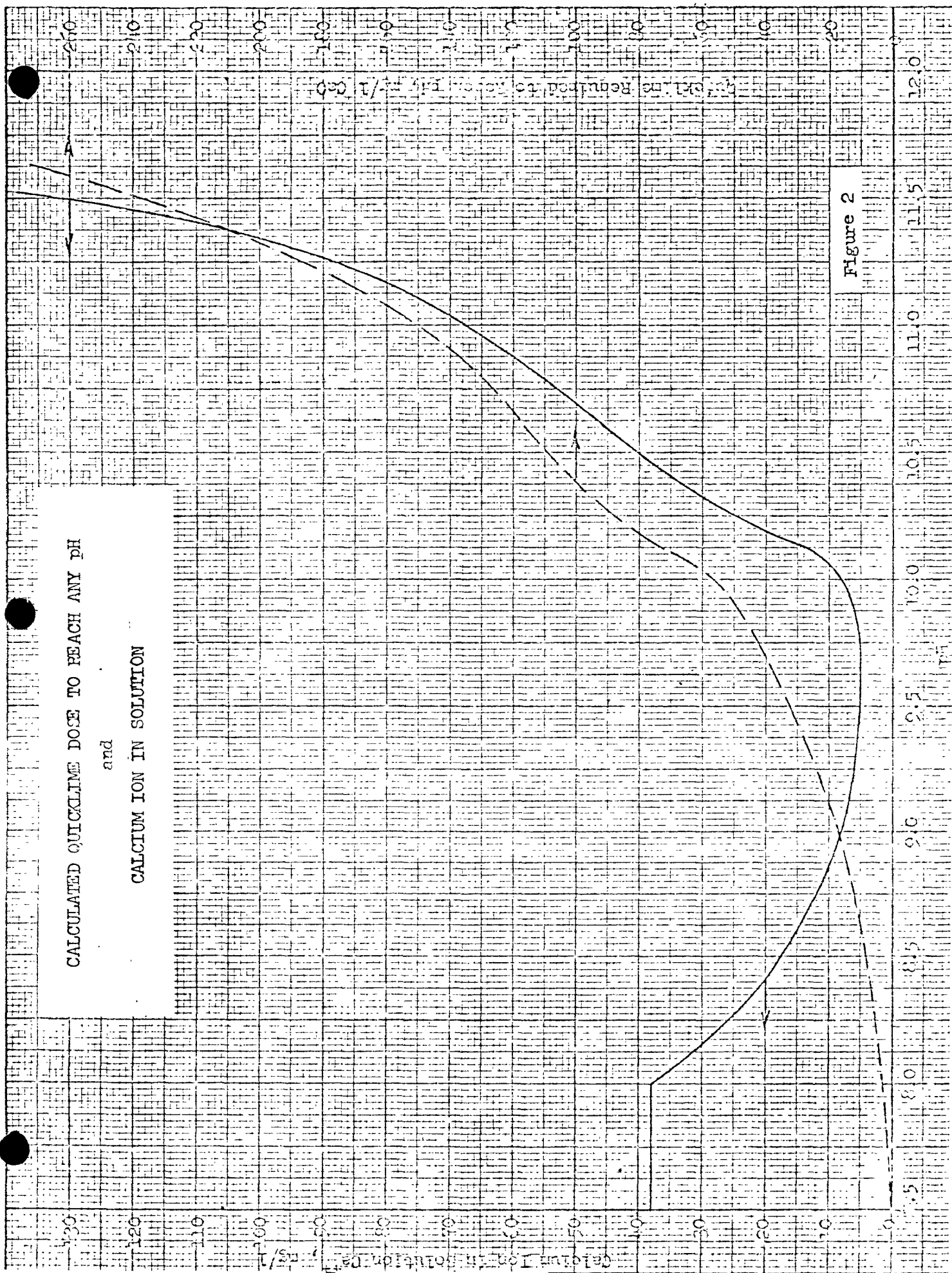
Figure 2

SOLIDS REMOVAL BY COAGULATION & SEDIMENTATION
Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity

Cost Adjusted to June, 1967
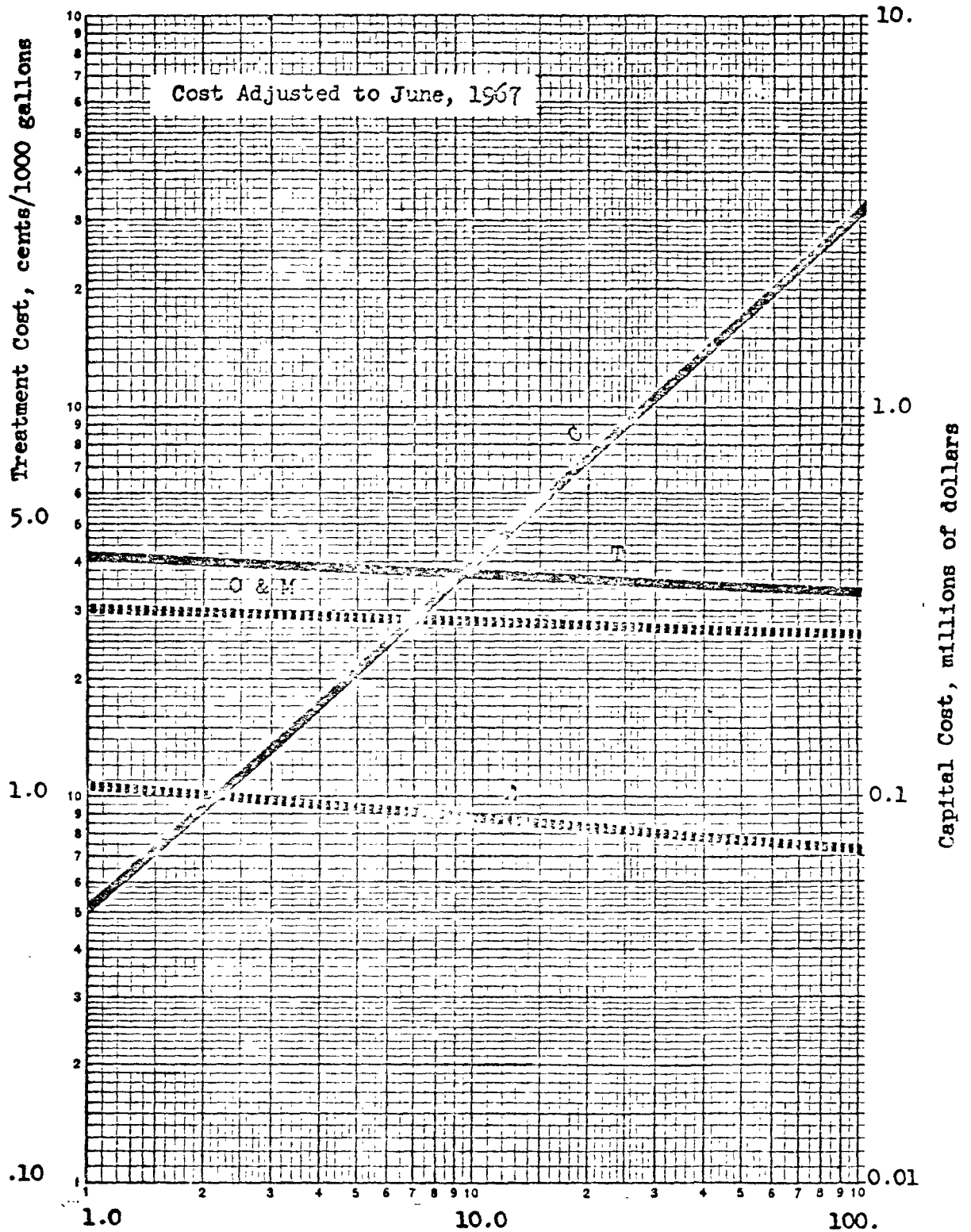
Design Capacity, millions of gallons per day

Figure 3

C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons(4 1/2 - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

PHOSPHATE REMOVAL FROM WASTE EFFLUENTS AND

RAW WASTES USING CHEMICAL TREATMENT*

By A. A. Kalinske[1] and G. L. Shell[2]

We have been conducting during the past several years various
pilot plant studies involving the removal of phosphates, and in
some cases improving effluents generally, using chemical treatment
in solids-contact units. Also, our company has built several such
plants (Reactor-Clarifiers) for the treatment of sewage treatment
plant effluents for industrial uses, primarily for power plant
cooling water. Two of these plants have been in operation for
about 3 and 6 years at Las Vegas, Nevada power plants. This is a
summary report in which data will be presented from the pilot plant
studies and, also, from one of the full-scale plants at Las Vegas,
for which fairly extensive operating data have been kept for over
2 years.

A. Pilot Plant Studies at Colorado Springs, Colorado
Of Trickling Filter Plant Effluent

The City of Colorado Springs has a conventional trickling
filter treatment plant which produces an effluent that is not of
particulary high quality, especially during the winter, as the
plant is somewhat overloaded. The City decided to embark on a
water reclamation program, and part of this was the treatment of
a portion of the effluent as make-up cooling water for their power
plant.

Among the requirements for this reclaimed effluent was that the total phosphate not exceed about 1.0 mg/l as P, and preferably less.

A cooperative arrangement was entered into between Eimco Corp. and the City to carry on various pilot plant studies which have now been completed. It was decided to use teritary chemical treatment of the effluent, using a solids-contact type unit, and to study both lime and alum treatment. The pilot plant was designed to handle about 25 gpm, which gave a hydraulic loading of 1400 gpd/ft^2 and a total retention time of 1.25 hrs.

The total test period occupied about 1½ years so all the variations in sewage effluent characteristics would be encountered, including the temperature variations, which had a pronounced effect on the quality of the STP effluent. The data presented herein are the averages of months of test work and, therefore, we believe can be used with confidence both from the technical and economic standpoints.

In Tables 1 and 2 are shown the phosphate removals obtained with various dosages of lime and of alum. All phosphorus values are expressed as total P. Note that these data are on unfiltered samples obtained from the solids-contact unit, which was operated at about an overflow rate of 1400 gpd/ft^2, though we would recommend that the lime treatment could be operated at about 1800 gpd/ft^2 while the alum treatment unit should not be operated over 1200 gpd/ft^2 due to the difference in settling characteristics of the floc and the slurry produced in the solids-contact unit.

Filtration of the samples through No. 42 Whatman paper further reduced the effluent P about 60%. The suspended solids from the treatment unit averaged about 10-20 mg/l.

In Tables 3 and 4 are shown a summary of what we considered, for the treatment desired, was the lowest lime and alum dosages that could be used, not only for obtaining the desired phosphate removal but for removal of other materials as indicated, since the solids-contact unit was to be followed by filtration and activated carbon contacting.

One could conclude that an 80% P removal could have been obtained with a lime dosage of about 200-250 mg/l, or an alum dosage of about 125-150 mg/l plus a samll amount of coagulant aid.

For the dosages of 325 mg/l of lime or 175 mg/l of alum (plus 1/4 mg/l of polymer aid), Table 5 shows an economic comparison for the 2 types of treatment, and the unfiltered effluent quality that can be expected. For the lime treatment the effluent was neutralized with acid, at a cost indicated, from a ph of 10.5 to about 7.0. The total chemical costs appear to be about 5 cents/ 1000 gals., for either treatment, and with essentially comparable effluent quality, though the alum treatment does give a lower BOD and COD.

The full-scale plant at Colorado Springs is to have a capacity of 2 MGD.

The solids-contact unit installed cost of this size will be about $75,000 which when amortized over 20 years at 5% amounts to about 1 cent/1000 gal. If reclamation of the lime from the sludge would be feasible for this size of plant, the operating cost could be reduced, but more capital equipment would be needed.

However, in evaluating the difference in total costs of the lime and alum treatments consideration must be given to the difference in the character of the sludge produced. On the basis of the above described tests there was 4000 lbs/day of dry solids produced per million gallons of effluent with the lime treatment at the dosage of 325 mg/l. This sludge thickened by gravity readily to 10% solids, which means 5000 gals/day of sludge for disposal. With the alum treatment, 2000 lbs/day of solids per million gals were produced, but the volume was 12,000 gals since it could only be thickened to 2% by weight by gravity. Such alum sludge could be mixed with the waste biological sludge and dewatered on a vacuum filter. The lime sludge could be dewatered on a vacuum filter, and at 5000 gals/day such a filter would cost, installed, about $8,00

B. Pilot Plant Studies Using Lime Treatment of Raw Sewage

It has been shown that phosphates can be removed by using chemical treatment on the raw sewage, and this might have overall economic advantages since such treatment would also remove a significant amount of the BOD, thus permitting reduction of the secondary biological plant, with resultant less biological sludge produced.

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The same solids-contact unit, described previously, was operated on screened raw sewage at Colorado Springs at a hydraulic loading of 1400 gpd/ft^2. The results obtained are shown in Table 6.

Note that the effluent P obtained was not quite as good, for the same lime dosage, as was obtained when treating the effluent. Thus for 300 mg/l of lime, the P was reduced on an average from 6.2 to 0.7 (89%), while with 325 mg/l added to the effluent the reduction was from 9.7 to 0.5 (95%). There is every reason to expect this, as undoubtedly a portion of the lime is "tied up" by reacting with the various complex organics present in raw sewage. These results are contrary to some others that have been reported, and we therefore checked this further in our Salt Lake City Laboratory using raw and treated SLC sewage, and obtained essentially the same general difference.

However, we definitely confirmed that solids recycle, and the contact with previously formed solids, as obtained in a solids-contact unit, is extremely beneficial in obtaining the full benefits of the lime used.

As we said initially, this method of treatment should be evaluated against teritary treatment of the effluent where it can be used, since even though somewhat more lime will be required, it does reduce the BOD load on the secondary plant. However, it cannot be used economically simply be adding the lime to the raw sewage entering an existing primary clarifier.

Good mixing and solid-contacting are essential, and this would require significant modifications to any existing "conventional" biological treatment plant of whatever type.

C. Phosphate Removal In Full-Scale Plants
At Las Vegas, Nevada

The Nevada Power Co. previous to 1961 had used the effluent from the Las Vegas trickling filter treatment plant for cooling water. However, as the phosphates gradually increased, averaging 35 mg/l as PO_4, it was mandatory that they be removed before the effluent could continue to be safely used. Therefore, in 1961 a 2500 gpm solids-contact unit (Eimco Reactor-Clarifier) was installed to remove phosphates and also excessive suspended solids from the sewage plant effluent. In 1964 another power station was built and a 2000 gpm solids-contact unit was installed at this plant. The necessary quantity of sewage effluent is pumped to these 2 treatment units. These installations with some operating data are described in a paper by W. H. Johnson, in the Proc. of the International Water Conference, Engrs. Soc. of Western Penna., Sept. 1964, p. 73.

The prime purpose of these treatment plants was to reduce the phosphates to low values. In Table 7 data are given, as 2-year averages, for one of the units. Note that P is reduced from about 10.0 to 0.6 at a lime dosage of 240 mg/l with 3 mg/l of a coagulant aid.

The chemical costs are about 5 cents per 1000 gals. The installed cost of each solids-contact unit, including piping, feeders, etc. was about $100,000, which amortized over 20 years at 5% amounts to a cost of about 1 cent/1000 gal.

The sludge is lagooned in the desert; it concentrates to about 11% in the Reactor-Clarifier units. The excellent treatment obtained, with a significant hardness reduction, which is difficult to achieve when treating sewage plant effluent with lime, is due to the very high solids that are carried in circulation, thus providing the necessary solids-contact treatment in the reaction zone.

TABLE 1.

PHOSPHATE REMOVAL FROM TRICKLING FILTER EFFLUENT WHEN TREATED WITH LIME

Lime Dosage mg/l	Influent		Effluent*		
	pH	P mg/l	pH	P mg/l	% Removal
250	7.4	10.0	9.8	1.07	89
300	7.5	10.0	10.3	.93	91
350	7.4	10.4	10.8	.60	94
400	7.4	10.7	11.0	.60	95

* Unfiltered sample

TABLE 2.

PHOSPHATE REMOVAL FROM TRICKLING FILTER EFFLUENT WHEN TREATED WITH ALUM

Alum Dosage mg/l	Influent P mg/l	Effluent* P mg/l	% Removal
175	5.9	1.3	78
175 + ¼ Coag. Aid	10.5	1.0	90
200	8.8	1.3	85
225	9.0	0.7	92

* Unfiltered sample

TABLE 3.

TRICKLING FILTER EFFLUENT TREATMENT
(325 mg/l Lime + ¼ mg/l of Coag. Aid)

	Influent*	Effluent	% Removal
BOD, mg/l	34	9	74
SS, mg/l	54	11	80
P, mg/l	9.7	0.5	95
ABS, mg/l	2.6	2.2	18
Total Alk. mg/l	198	200	
Hard. mg/l	176	170	
pH	7.2	10.5	

* STP Influent

TABLE 4.

TRICKLING FILTER EFFLUENT TREATMENT
(Alum 175 mg/l + ¼ mg/l of Coag. Aid)

	Influent*	Effluent	% Removal
BOD, mg/l	34	5	85
SS, mg/l	37	11	70
P, mg/l	10.7	1	91
ABS, mg/l	2.3	1.5	35
pH	7.1	6.7	

* STP Effluent

TABLE 5.

COMPARISON BETWEEN LIME AND ALUM TREATMENT IN SOLIDS-CONTACT UNIT OF TRICKLING FILTER EFFLUENT

CHEMICALS	Alum (1)	Lime (2)
Coagulant, mg/l	175	325
Coag. Aid, mg/l (4)	¼	¼
Neutralizing Acid, lb/MG (3)	—	1340
Total Cost, $/MG	52	55
EFFLUENT QUALITY		
BOD, mg/l	5	9
COD, mg/l	47	88
SS, mg/l	11	11
Total P, mg/l	1	0.5

(1) Based on 4 cents/lb.
(2) Based on 1.25 cents/lb.
(3) Based on 1.1 cents/lb.
(4) Based on $2.00/lb.

TABLE 6.

TREAMENT OF RAW SEWAGE WITH LIME IN A SOLIDS-CONTACT UNIT

| BOD, mg/l | | | COD, mg/l | | | S.S., mg/l | | | Total P, mg/l | | | pH | | Treatment |
Inf.	Eff.	% Rem.	Inf.	Eff.	% Rem.	Inf.	Eff.	% Rem.	Inf.	Eff.	% Rem.	Inf.	Eff.	
26	118	48	439	231	47	175	74	58	8.0	2.8	65	7.1	9.3	100 mg/l
			430	192	55	224	114	49	8.3	2.3	72	7.3	9.6	200 mg/l
21	90	57	394	171	57	174	65	63	6.2	0.7	89	7.4	11.1	300 mg/l
			426	132	69	188	53	72	7.4	0.7	91	6.9	11.2	400 mg/l
47	37	75	530	202	62	200	97	52	6.7	2.7	60	7.2	9.6	100 mg/l lime + ¼ mg/l Coag. Aid

TABLE 7.

TREATMENT OF TRICKLING FILTER EFFLUENT (1 MGD) IN SOLIDS-CONTACT UNIT: AVE. DATA FOR 1967 AND 1968.

(Lime Dosage 240 mg/l + 3 mg/l Nalco 603)

	Influent	Effluent
Total P, mg/l	10.0	0.6
BOD, mg/l	21	
COD, mg/l	88	
SS, mg/l	31	5
TDS, mg/l	755	630
Hardness, mg/l	352	298
Calcium, mg/l	155	190
Magnesium, mg/l	192	118
Total Alk., mg/l	285	132
pH	7.5	10.2

HIGH RATE SEDIMENTATION IN WATER TREATMENT WORKS

By

Gordon Culp, Research Manager
Sigurd Hansen, Research Engineer
Gordon Richardson, Laboratory Manager

Neptune MicroFLOC, Inc.
Corvallis, Oregon

Presented on June 4, 1968, at the American Water Works Association's 88th Annual Conference, Cleveland, Ohio.

PRINCIPLES OF HIGH RATE SEDIMENTATION

It has long been recognized that a settling basin should be as shallow as possible and that detention times of only a few minutes could be used in very shallow basins. For example, a particle settling at a rate of 1 inch per minute requires 120 minutes to fall to the bottom of a conventional clarifier of 10 foot depth. If the basin were 2 inches deep, this particle would fall to the bottom in only two minutes. In 1904, Hazen (1) presented his argument that settling basin efficiency is primarily dependent upon basin depth and overflow rate and is independent of detention time. He proposed basin depths of as little as 1 inch. Over 20 years ago, Camp (2) proposed settling basin depths of 6 inches and total settling basin detention times of 10 minutes.

A detailed literature review by the authors (3) showed that there have been many attempts at applying the shallow depth sedimentation principles proposed by Hazen and by Camp. Wide, shallow trays were generally inserted within basins of conventional design. These attempts met with only limited success because of two major problems: (a) the unstable hydraulic conditions encountered with very wide, shallow trays, and (b) the minimum tray spacing was limited by the vertical clearance required for mechanical sludge removal equipment. The authors have overcome both of these problems by using very small diameter tubes rather than wide, shallow trays. Longitudinal flow through tubes with a diameter of a few inches offers theoretically optimum hydraulic conditions for sedimentation and overcomes the hydraulic problems associated with tray settling basins. Such tubes have a large wetted perimeter relative to the wetted area and thereby provide laminar flow conditions as evidenced by very low Reynolds numbers. Fischerstrom (4) felt Reynolds number of less than 500 in settling basins would be most beneficial to the settling process. A 1 inch diameter tube, 4 feet long, through which water is passed at a rate of 10 gpm per square foot of cross-sectional area has a Reynolds number of only 24 while providing an equivalent surface overflow rate of 235 gpd per square foot. The 3 minute detention time of such a tube settling device under these conditions certainly makes the cost and space saving potential apparent. The authors now have tube settling devices in operation in many water treatment plants which are providing excellent clarification with settling detention times of <u>less than 10 minutes</u>.

The authors recently made detailed presentations of their research on techniques for applying shallow depth tubes as sedimentation devices (5). The purpose of this paper is to present operating experiences with applications in water treatment plants and to present additional research data.

BASIC TUBE SETTLER CONFIGURATIONS

The authors have described (3, 5) two basic tube configurations which are shown in Figure 1: (a) essentially horizontal and (b) steeply inclined. The operation of the essentially horizontal tube settlers [described in detail in reference (3)] is coordinated with that of the filter used to clarify the tube settler effluent. Each time the filter backwashes, the tube settler is completely drained. The falling water surface scours the sludge deposits from the tubes and carries them to waste. The water drained from the tubes is replaced with the last portion of the filter backwash water. The tubes are inclined only slightly in the direction of flow (5°) to promote the drainage of sludge during the backwash cycle. If the inclination of the tubes is increased to a steep angle (45° - 60°), continuous gravity drainage of the settleable material from the tubes can be achieved (5). The incoming solids settle to the tube bottom and then exit the tubes by sliding downward along the tube bottom. A flow pattern is established in which the solids settling to the tube bottom are trapped in a downward flowing stream of concentrated solids. This countercurrent flow of solids aids in agglomerating particles into larger, heavier particles which settle against the velocity of the upwardly flowing liquid. The continuous sludge removal achieved in these steeply inclined tubes eliminates the need for drainage or backflushing of the tubes for sludge removal.

ESSENTIALLY HORIZONTAL TUBE SETTLER

Applications

The essentially horizontal tube settler has been used primarily in small (15,000 gpd) to medium sized (10 MGD) water treatment plants where the elimination of operator attention for sludge removal from the clarifier is a significant benefit. By draining the tubes each time the filter backwashes, positive sludge withdrawal from the clarifier is achieved. The entire backwash - tube drainage cycle is automated. Thus, no operator judgement on when or how much sludge to withdraw from the clarifier is required. A schematic of a package water treatment in which tube settling has been used is shown in Figure 2. The detention time within the tubes, at design flow, is 6 minutes. The impact of this low residence time on the plant dimensions is well illustrated by the 6 foot high by 6 foot wide by 14 foot length dimensions for a complete 100 gpm water treatment plant providing flocculation, sedimentation and filtration in one rectangular, prefabricated, steel plant. As shown in Figure 2, the coagulated raw water may first be passed upwards through a fluidized calcite column which automatically maintains the pH in the proper range for good coagulation. If alum is overfed, the calcite will automatically buffer the coagulated water. The use of calcite for this purpose eliminates one possible cause (overdosage of coagulant) of poor finished water turbidity.

The tubes used in these essentially horizontal tube settlers are hexagonal in shape. The hexagonal tubes nest together to form a honeycomb pattern, as shown in Figure 3. The tremendous wetted perimeter of such a tube configuration relative to its wetted area provides extremely low Reynolds numbers, well within the laminar flow range.

The effects of tube length, diameter, and flow rate on tube settler efficiency have been presented earlier (3). Tube diameters of 1 to 2 inches and lengths of 2 to 4 feet are used in most water treatment applications. Hydraulic loading rates of 3 to 5 gpm per square foot of tube entrance area are generally used. Data have been presented (3) which show that the tube settler, mixed-media filter combination in a plant with total detention time of less than 30 minutes provides efficient clarification of very turbid waters (1000 JU), highly colored waters, waters containing filter-clogging algae, waters containing iron and manganese, and raw waters with taste and odor. Since these data were published, confirming operating data from several plants with capacities of 30,000 gpd to 3 MGD have been obtained from field installations. The partial list of these plants shown in Table I illustrate the wide range of water quality being subjected to treatment in plants employing the basic flow pattern shown in Figure 2. The majority of these plants are being used for potable supplies and are operating at total plant detention times of about 30 minutes. Although this detention time is greatly less than that in plants employing conventional sedimentation techniques, data collected in prototype studies indicate the plants are actually conservatively rated. To illustrate this, Figure 4 presents data collected during one of these studies with a plant as shown in Figure 2. During this test, each of the plant components was being operated at rates considerably higher than the design criteria normally used for these plants, i.e., tube settler detention 3 minutes rather than 6 minutes, filter rate 8.5 gpm per square foot rather than 5 gpm per square foot, flocculation time 5.4 minutes rather than 10 minutes. The plant operating with an overall detention time of 16 minutes reduced average raw water turbidity of 1000 JU to an average of less than 0.1 JU. Although the filter rate of 8.5 gpm per square foot resulted in a relatively high initial headloss (2.7 feet H20), the percentage backwash water at the end of the 8 hour run was only 2.5 percent. If the run had been continued to the normal backwash headloss value of 8 feet, the backwash requirement would have been less than 2 percent. These data well illustrate how the mixed-media filter (coal, sand, garnet) complements the tube settler to accomplish what may certainly be classed as "high rate" clarification.

Flow Distribution

Flow distribution problems are much less severe in the tube settler than in earlier tray settling devices. One of the major reasons is the extremely stable hydraulic condition established within the tubes. As discussed earlier in this paper, laminar flow conditions are established in the tubes. Thus, there are no turbulent flow conditions to promote short circuiting. Also, it has been found that the sludge deposits within the tubes act as flow distribution aides. If one tube is receiving more flow than another, the more rapid buildup of sludge in the first tube will cause some flow to be diverted to the second tube. The sludge deposits themselves thus act as a "self-orificing" device in the horizontal tube settlers. Of course, care must be taken in the design of the tube inlet and outlet conditions so that no great velocity gradients are established across either the inlet or outlet faces of the tube modules.

4

TABLE I

Partial List of Installations of Water Treatment Plants
Using Horizontal Tube Settlers and
Mixed-Media Filtration

Location	Facility Served	Plant Capacity (gpm)	Treatment Problem
Alabama	Paper Mill	100	10-100 JTU Turbidity
Ohio	Subdivision	20	10-65 JTU Turbidity
Alabama	Nuclear Reactor	100	2-28 JTU Turbidity
Tennessee	Recreational Area	20	2-30 JTU Turbidity
Oregon	Recreational Area	20	30-150 JTU Turbidity 20 Color
Venezuela	School	20	10-40 JTU Turbidity
Manitoba	Municipality	60	10-35 JTU Turbidity
Pennsylvania	Power Station	100	5-15 JTU Turbidity, pH 3.5, Iron 2.8 mg/1, Manganese 1.0 mg/1
Wyoming	Oil Field Reuse	200	Oil and Suspended Solids
Mississippi	Municipality	2,000	3-5 mg/1 Iron
Massachusetts	Municipality-Pilot	20	200 Color
New Mexico	Recreational Area	20	10-20 JTU Turbidity
West Virginia	Municipality	350	10 JTU Turbidity
British Columbia	Municipality	350	4.5 mg/1 Iron, 160 Color
Pennsylvania	Power Station	100	25 JTU Turbidity, 1 mg/1 Iron, 20 Color
Idaho	Municipality	100	10-1000 JTU Turbidity, 20 Color
Oregon	Municipality	500	5-10 JTU Turbidity 20 Color, Algae

Flow distribution analyses have been made in a 20 gpm plant as shown in Figure 2, utilizing the salt tracer technique (6). A batch addition of a solution containing 50 grams per liter of sodium chloride was dispersed into a stabilized flow of untreated source water in the flocculator.

Conductivity analyses of samples collected at five minute intervals from the influent and twelve vertical and lateral effluent settler locations showed a variation in peak value of ±10 micromhos on triplicate runs. Comparing the variation in conductivity with the corresponding salt as chloride concentration indicated sufficient linearity existed to allow the data to be evaluated on the basis of conductivity.

Distribution of the inlet flow to the individual tubes can be considered satisfactory based upon the minor variation experienced in peak effluent conductivity and that the individual tube samplings were found to have retention times 4 percent less than the theoretical sampling retention period.

Analyses of the tube effluent composite sample with time indicated a minimum of short circuiting existed in that the volumetric displacement efficiency was found to be 84 percent as compared with that of 63 percent listed (6) for ideal basins.

STEEPLY INCLINED TUBE SETTLERS

As discussed earlier, the solids which settle to the bottom of a tube inclined at a steep angle (greater than 45°) will slide down the tube bottom continuously. This enables sludge removal to be achieved without draining or backflushing the tubes.

Although the benefits of this continual sludge removal phenomenon are obvious, the effects of steeply inclining the tubes on the path of the particles as they settle requires more detailed consideration. The path traced by a particle settling in a tube is the resultant of two vectors: V, the velocity of flow through the tube and v_S, the settling velocity of the particle. It can be seen in Figure 5 that if the settling surfaces are inclined upward in the direction of flow, the settling path of the particle is altered because the component of the settling velocity which is parallel to the tube wall, v_{sh}, is opposite in direction to the velocity vector V. If V is greater than v_S, the required length of the settling surface decreases as the angle increases from zero up to about 25 to 30 degrees (at $V=2.5\ v_S$) and then increases, approaching infinity as the angle of inclinitation is increased to 90 degrees. For $V<v_S$ the tray length continues to decrease with increasing angle.

The research data on performance of steeply inclined tube settlers in waste water treatment applications has been published (5). The following sections of this paper present research and field data on water treatment applications.

Laboratory Studies

During a continuation of the research studies earlier published (3), an apparatus was designed to study the effects of tube inclination on settling efficiency (See Figure 6). It was during the operation of this equipment that the "self-cleaning" phenomenon was first observed. Initial tests were carried out with five individual tubes inclined at angles of 0, 5, 20, 45, and 90 degrees. The sludge settling to the bottom of the tube inclined at 45° was observed to be continously moving downward and eventually falling into the inlet plenum. Some of the data collected on tube settling efficiency at the various angles of inclination are summarized in Figure 7. It was noted that tube efficiency showed an increase as the angle of inclination was increased to 35°-45° and then began to decrease the angle of inclination was increased further. However, results comparable to those obtained at 5° inclination were achieved at angles as steep as 60°. It appeared that as the angle of inclination was increased to the point where the settled sludge began to move down the tube bottom, that additional flocculation occurred as the heavier floc settled and collided with the smaller, upward moving floc, contributing to the increased efficiency over that achieved at 5°. However, a continuing increase in angle eventually results in the tube acting as an upflow clarifier and the advantages of the shallow tube depth are lost, resulting in a decrease in efficiency.

Following observation of this self-cleaning principle, the apparatus shown in Figure 6 was modified to better define the effects of various inclinations on sludge cleaning and on sedimentation efficiency. The

TABLE II

Tube Settler Efficiency in Preliminary Field Tests

(Tube Length – 2 Feet, Tube Diameter = 1.5 Inches)

Hydraulic Loading, Settling Tubes (gpm/ft2 of end area)	Polyelectrolyte Dose mg/l*	Aver. Raw Turbidity JU	Aver. Tube Settler Eff. Turbidity JU	Change-in-Filter Headloss, in. Water/Hr.	Filter Rate (gpm/ft2)
		TUBES INCLINED AT 5°			
3.7	0	250	37	1.0	6.3
3.7	0.1	250	17	3.0	6.3
4.0	0.2	250	70	1.75	6.8
4.0	0.5	230	16	0.5	6.8
5.0	0.2	250	21	4.0	8.5
		TUBES INCLINED AT 60°			
5.0	0.2	260	6	0.8	8.5
5.0	0	290	26	0.8	8.5
6.75	0	270	45	0.85	8.5
6.75	0.2	240	13	2.3	8.5
8.5	0	250	45	2.0	10
8.5	0.2	250	14	1.0	10

*Added ahead of flocculator

tubes were repositioned at angles of 35, 40, 45, and 60 degrees. A slight decrease in efficiency (Figure 8) was noted as the angle of inclination approached 60°. However, the self-cleaning action was enhanced as the angle was increased from 45° to 60°. To insure adequate sludge removal from the tubes, an angle of inclination of 60° was used in the subsequent tests of multi-tube units.

Field Evaluation - Pilot Plant Scale

A plant of the type shown in Figure 2 was modified for the first field evaluations of the steeply inclined tubes. As shown in Figure 9, the plant was evaluated with the tubes at a 5° inclination and at a 60° inclination. The tubes used were 2 feet in length and 1-1/2 inch in diameter. Because the tube chamber was originally designed for 4 foot long tubes inclined at 5°, a portion of it was blocked off by the plywood diaphragm shown in Figure 9 for the tests with 2 foot tubes inclined at 5°. The tubes were installed so that the inlet and outlet conditions and the total tube entrance area were the same with both the 5° and 60° tubes. The same mixed-media (coal, sand, garnet) filter was used to filter the tube effluent in both cases. The surface water being treated was coagulated with alum and, as noted in Table II, polyelectrolyte was added in some cases. The data shown in Table II show that the water quality produced by the 60° tubes at 8.5 gpm per square foot was lower in turbidity than that produced by the 5° tubes at 5.0 gpm per square foot, with 0.2 mg/l polyelectrolyte used in both cases. The tube effluent quality was compatible with the mixed-media filter in all cases and filter runs to 8 feet of headloss were greater than 18 hours, in all cases. Data collected during one run of the 60° tubes is shown in Figure 10. The effect of polyelectrolyte on tube settler efficiency is clearly shown by the sudden decrease in tube effluent turbidity following the beginning of polyelectrolyte feed at 4.2 hours. The filter effluent turbidity remained less than 0.1 JU throughout the run. The tube settler detention time under the conditions shown in Figure 10 was 2.3 minutes.

Field Evaluation of Modular Tube Units - Pilot Plant Scale

Because of the very encouraging results obtained in the preliminary field tests described above, work was begun on the design of a modular unit of steeply inclined tubes which would minimize installation problems. Following preliminary evaluation of a great many potential designs, this development work resulted in the tube module design shown in Figure II (patent pending) in which the material of construction is normally PVC. Extruded PVC channels are installed at a 60° inclination between thin sheets of PVC. By inclining the tube passageways, rather than inclining the entire module, the rectangular module can be readily mounted in either rectangular or circular basins. By alternating the direction of inclination of each row of the channels forming the tube passageways, the module becomes a self-supporting beam which needs support only at its ends. Following the development of this module, field tests of its efficiency as a sedimentation device were begun. A tube cross-section of 2 inches by 2 inches and a tube length of 24 inches was used in the following tests.

The apparatus (shown in Figure 12) was set up at the authors' laboratory. The laboratory ground water supply was used. A mud slurry was mixed with the incoming water to provide various levels of raw water turbidity. Alum (40 mg/l) was added as the primary coagulant with polyelectrolyte additions made in some tests. Tube loading of 4-6 gpm per square feet were investigated (tube entrance area = 9 ft.^{2}) with raw water turbidities of 50 and 250 JU. The data from these tests are summarized in Table III. In some runs, as noted, the flocculator drive motor was turned off to evaluate the tube efficiency without prior mechanical flocculation.

At the lower rate of 4 gpm per square foot, the addition of polyelectrolyte did not markedly improve the effluent clarity. However, when the flow rate was increased to 6 gpm per square foot, the higher settling velocities imparted by the polyelectrolyte were of significant benefit. When the flocculator was operated, the turbidities were fairly constant throughout the run. However, when the flocculator motor was not operated, it was found that the effluent turbidity decreased with time as the solids concentration beneath the tubes increased. This is not surprising since solids contact in and beneath the tubes was the prime source of flocculation in this case. Although it was found that the sludge blanket could be established with the steeply inclined tube settler without subjecting the incoming water to mechanical flocculation, flocculation hastened the development of the blanket. After the sludge blanket was well established, the flocculator could be turned off with no noticeable effect on the clarified effluent quality. This observation suggests that by maintaining an upflow of newly coagulated water through a region of high solids concentration, the external flocculation requirements can be significantly reduced. This principle, of course, is recognized and capitalized on by solids contact clarifier manufacturers.

These tests indicated the steeply inclined tube modules shown in Figure II performed well as a sedimentation device and were capable of producing settled water turbidities consistent with the capabilities of the mixed-media filter under all the conditions shown in Table III.

Field Evaluation - Plant Scale

The next logical step in the development of the steeply inclined tube settling process was a plant scale application. Fortunately, the city of Newport, Oregon, and their consulting engineer were faced with a water treatment plant expansion at the time when the tube settling experiments described above were being completed. The existing 1.5 MGD plant (Figure 13) consisted of a circular flocculator-clarifier followed by rapid sand filters. The raw water characteristics are as follows: turbidity - 10 to 20 JU; color - 50 to 130 units; pH - 7.6; iron - 4.7 mg/l; temperature - 66° F; alkalinity - 85 mg/l; and hardness - 17 mg/l. The plant operator normally applies about 35 mg/l alum and several mg/l of activated carbon to the raw water to produce an acceptable finished water quality. Pilot tests were conducted using a plant of the type shown in Figure 2. It was found that an alum dose of 30 mg/l polyelectrolyte feed of 0.3 mg/l, and 1.5 mg/l chlorine would enable the tube settler - mixed-media filter combination to produce a finished water quality of 0.15 JU turbidity, 5 color units, and 0.1 mg/l iron. The tube settler and mixed-media filter were both operated at 5 gpm per square foot in these pilot tests.

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TABLE III

Performance of Pilot Plant with Steeply Inclined Tube Modules

Flow Rate gpm/ft^2	Floc Time Min.	Polyelectrolyte Dosage mg/l	Average Raw Turbidity JU	Average Settled Turbidity JU
4	7	0	50	18
4	7*	0	92	20
4	7*	0.1	50	20
4	7*	0.2	54	13
6	4.5	0.1	53	27
6	4.5	0.2	52	21
6	4.5	0.2	231	27
6	4.5*	0.2	246	54
6	4.5*	0	49	35
6	4.5	0	255	52

*Flocculator drive motor not operated

Based upon the pilot test results, modification of the full scale plant
was begun late in the fall of 1967 in order to increase the plant capacity
from 1.5 MGD to 3.0 MGD by installation of tube modules in the existing
clarifier and by conversion of the rapid sand filters to mixed-media
beds. As a first step, tube modules were to be installed in the existing
clarifier to evaluate their performance on a plant scale.

Because the available water supply to the clarifier was to be limited by
the existing 1.5 MGD raw water pump during the early tests, tubes were
installed in only a portion of the basin. Tube modules of the type
shown in Figure II were used. The tubes were installed over 1/6 of the
clarifier surface as shown in Figure 15. Tube modules were also used
as support beams for the upper modules, see Figures 15 and 16, so that
none of the clarifier surface was lost due to a support structure. These
radial support beams were attached by PVC pipe to support brackets on
the inlet well on one end and the effluent weir at the other end. The
supporting brackets, pipe, and modules are pictured in Figure 17 while the
installation of a support module is shown in Figure 18. Once the support
beams were in place, the remaining tube settler modules were placed in
position. The portion of the basin in which tubes were installed (pictured
in Figure 19) was isolated from the remaining part of the clarifier by a
plastic barrier attached radially to the clarifier on each side of the tube
section. Flow through the tube section was regulated by closing off portions
of the effluent weir in the rest of the basin by a galvanized sheet steel
plate clamped to the weir plate. Preliminary tests quickly led to closing off
this entire weir area and passing the entire plant flow through only the
210 square feet of the basin covered with tubes. Although the nominal
plant flow was 1.5 MGD, flow measurement during the test period indicated
the raw water pump was actually delivering only 910 gpm. Thus, the hydraulic
loading on the tube area was 4.3 gpm per square foot.

Flow distribution analyses and effluent quality determinations were made.
Because only the existing peripheral effluent weir was used, less-than-
perfect flow distribtuion was anticipated. Although radial collection
weirs would greatly aid in flow distribution, it was desired to first
evaluate the performance using only the existing weir.

The rise rate of hydrochloric acid injected into each tube module at
several points was used to determine the velocity in each module. For
the purposes of identification, the modules were labeled "A," "B," "C,"
and "D" as shown in Figure 15. The resulting flow distribution data
are shown in Figure 20. As was expected, the outer modules nearest the
effluent weir were receiving the bulk of the flow and were operating
at 6.6 gpm per square foot as compared to the average of 4.3 gpm per
square foot based upon the entire surface area covered by tubes. Even
with this flow distribution, the tubes were performing as efficient
sedimentation devices. As shown in Figure 20, the tube effluent tur-
bidity increased only slightly as the flow rate increased from 2 gpm
per square foot in module D to 6.6 gpm per square foot in module A.
The tube effluent contained no settleable solids while samples collected
earlier from the existing clarifier indicated that its effluent frequently
contained 0.2 - 1.0 ml/l settleable solids and an average turbidity of
5.1 JU. The fact that the tube modules operating at 4.3 gpm per square
foot (average) were producing better effluent than the clarifier previously

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did operating at 0.7 gpm per square foot was further confirmed by the fact that the length of filter runs increased from 26 hours to 60 hours following the modification of the clarifier.

At the time of this writing, the final Newport plant conversion is being made. The filter media conversion is underway with the design rate for the mixed-media filter being 5 gpm per square foot. The final clarifier conversion is being made with a ring of tube modules being installed completely around the periphery of the basin. The tube ring will operate at a rate of 5 gpm per square foot at 3 MGD. The peripheral location of the modules was selected to take advantage of the existing effluent collection system and the resulting flow distribution. If more of the clarifier surface is eventually covered with tubes to further increase the basin capacity, additional effluent collection weirs to better distribute the flow would be needed to realize the full advantage of the additional tube modules.

Summary

Shallow tubes are very efficient sedimentation devices. Two basic tube configurations have been used: (a) essentially horizontal and (b) steeply inclined. Sludge is removed from the essentially horizontal tubes by automatically draining them each time the filter backwashes and refilling them with filter backwash water. Over 20 water treatment plants employing these horizontal tubes with sedimentation detention times of less than 10 minutes with capacities of 20 gpm to 2,000 gpm are now in operation. In tests described in this paper, a plant providing flocculation, tube sedimentation, and mixed-media filtration, produced potable water (0.1 JU turbidity) from a raw water turbidity of 1,000 JU with an overall plant detention time of 16 minutes. Flow distribution analyses show the shallow horizontal tubes enable good flow distribution to be readily achieved.

The continuous self-cleaning of sludge from tubes inclined at a steep angle allows sludge removal to be achieved without the need for draining the tubes. Laboratory and field tests show that an angle of 60° provides continuous sludge removal while still allowing the tube to function as an efficient sedimentation device. Pilot plant tests have shown these steeply inclined tubes to efficiently remove alum floc at rates as high as 8.5 gpm per square foot. These tests led to the development of tube modules which were installed in an existing clarifier to increase its capacity from 1.5 MGD to 3.0 MGD. Analyses of the full scale installation showed good clarification at rates of 6.6 gpm per square foot. The installation of the tube modules in an existing clarifier and the conversion of the filter to a mixed-media bed provides plant expansion with substantial savings in cost and space. The coupling of tube settlers and mixed-media filter allows a reduction in the size and cost of new treatment facilities. This combination provides new design concepts to achieve efficient treatment plant design to produce a given quality finished water from a given raw water or waste water.

Acknowledgements

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6. "Operation and Control of Water Treatment Processes." Cox, C. R. Monograph Series No. 49. World Health Organization, Geneva, Switzerland (1964).
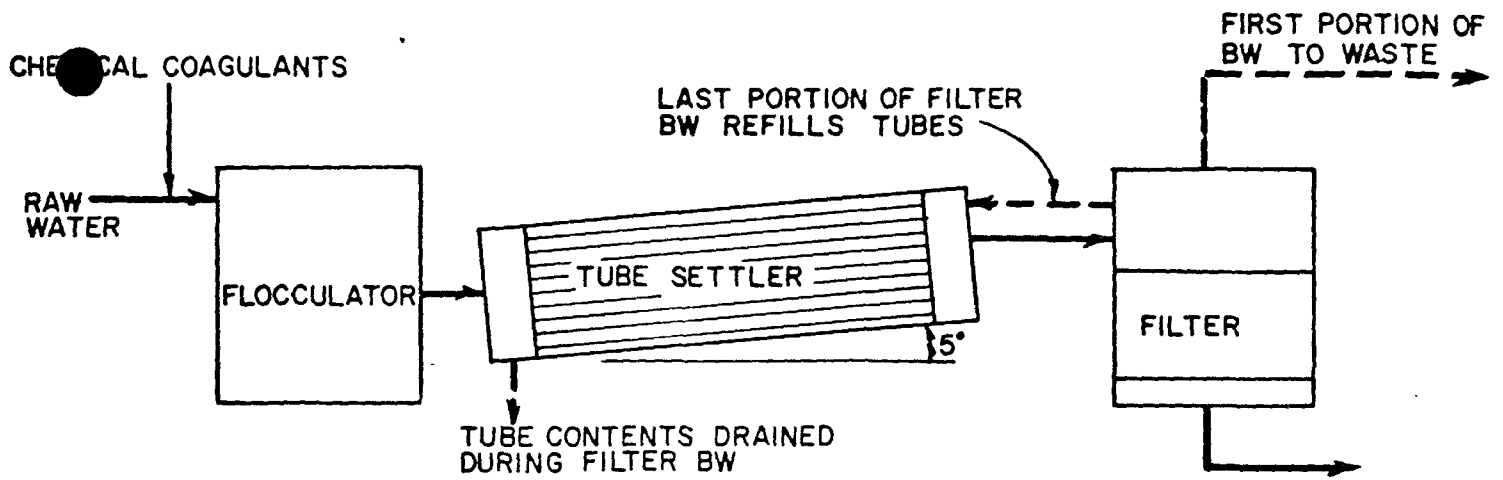
(a) ESSENTIALLY HORIZONTAL TUBE SETTLER
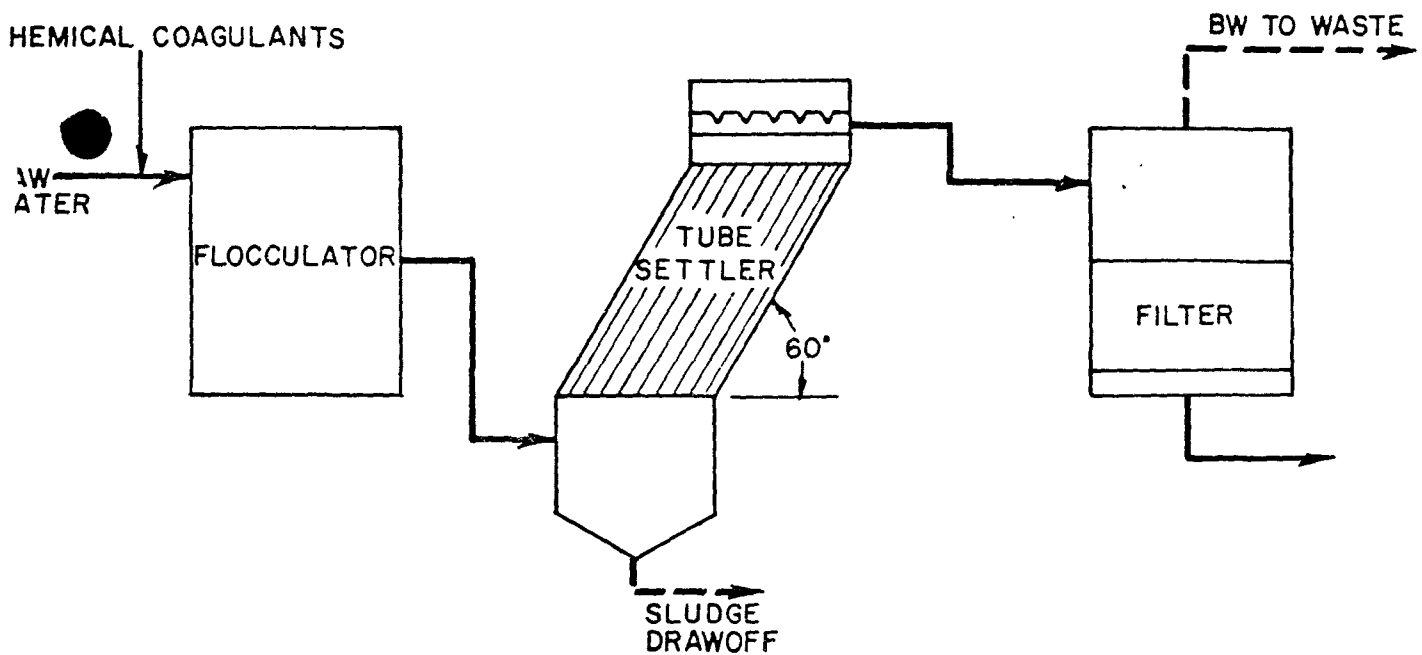
(b) STEEPLY INCLINED TUBE SETTLER

FIGURE I

BASIC TUBE SETTLER CONFIGURATIONS
SHOWN SCHEMATICALLY
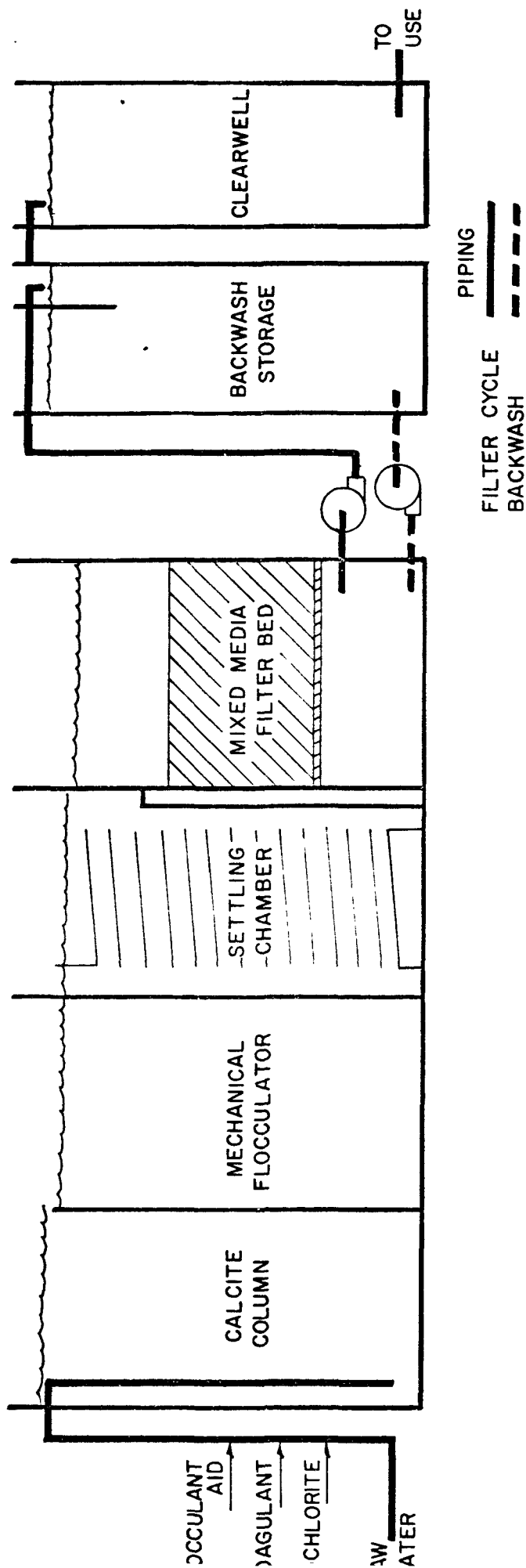
FIGURE 2

SCHEMATIC OF PACKAGE WATER TREATMENT PLANT

IN WHICH TUBE SETTLING IS USED

FIGURE 3

TYPICAL TUBE SETTLER MODULE

FOR USE IN PLANTS SHOWN IN FIGURE 2

HEADLOSS, FEET OF WATER

PLANT FLOW RATE — 34 G.P.M.
SETTLER FLOW RATE — 5 G.P.M./SQ FT
FILTER FLOW RATE — 85 G.P.M/SQ FT
ALUM DOSE — 50 PPM
POLYELECTROLYTE DOSE — 02 PPM
RAW WATER TURB — ~1000
TUBE SIZE — 2 FT LONG, 2 IN DIA
FLOCCULATION TIME — 5 4 MIN
TUBE SETTLER
    DETENTION TIME — 3 MIN
OVERALL PLANT
    DETENTION TIME — 16 MIN.

TUBE SETTLER
EFFLUENT TURBIDITY

FILTER BED HEADLOSS

FILTER EFFLUENT TURBIDITY

OPERATING TIME, HOURS

FIGURE 4

OPERATIONAL DATA FROM TREATMENT PLANT

SHOWN IN FIGURE 2
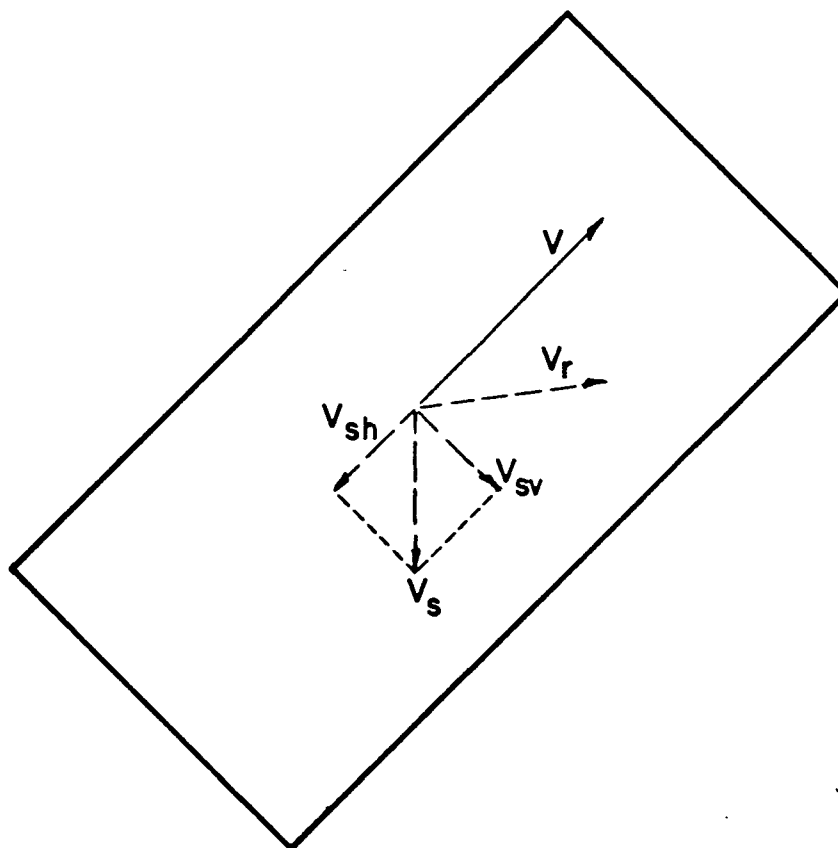
FIGURE 5

EFFECT OF TUBE INCLINATION ON SETTLING PATH OF
DISCRETE PARTICLE

FLOCCULATED WATER

INLET PLENUM

PLAN VIEW

FLOW CONTROL VALVE

1" DIA. x 4' LONG TUBES

TO DRAIN

FLEXIBLE PLASTIC
TUBING FOR
CONNECTION

2 VERTICAL
ENTRY PORTS

INLET
PLENUM

FLOCCULATED
WATER

3 HORIZONTAL ENTRY PORTS

MAG-MIXER USED TO PREVENT
DEPOSITION IN INLET PLENUM

ELEVATION
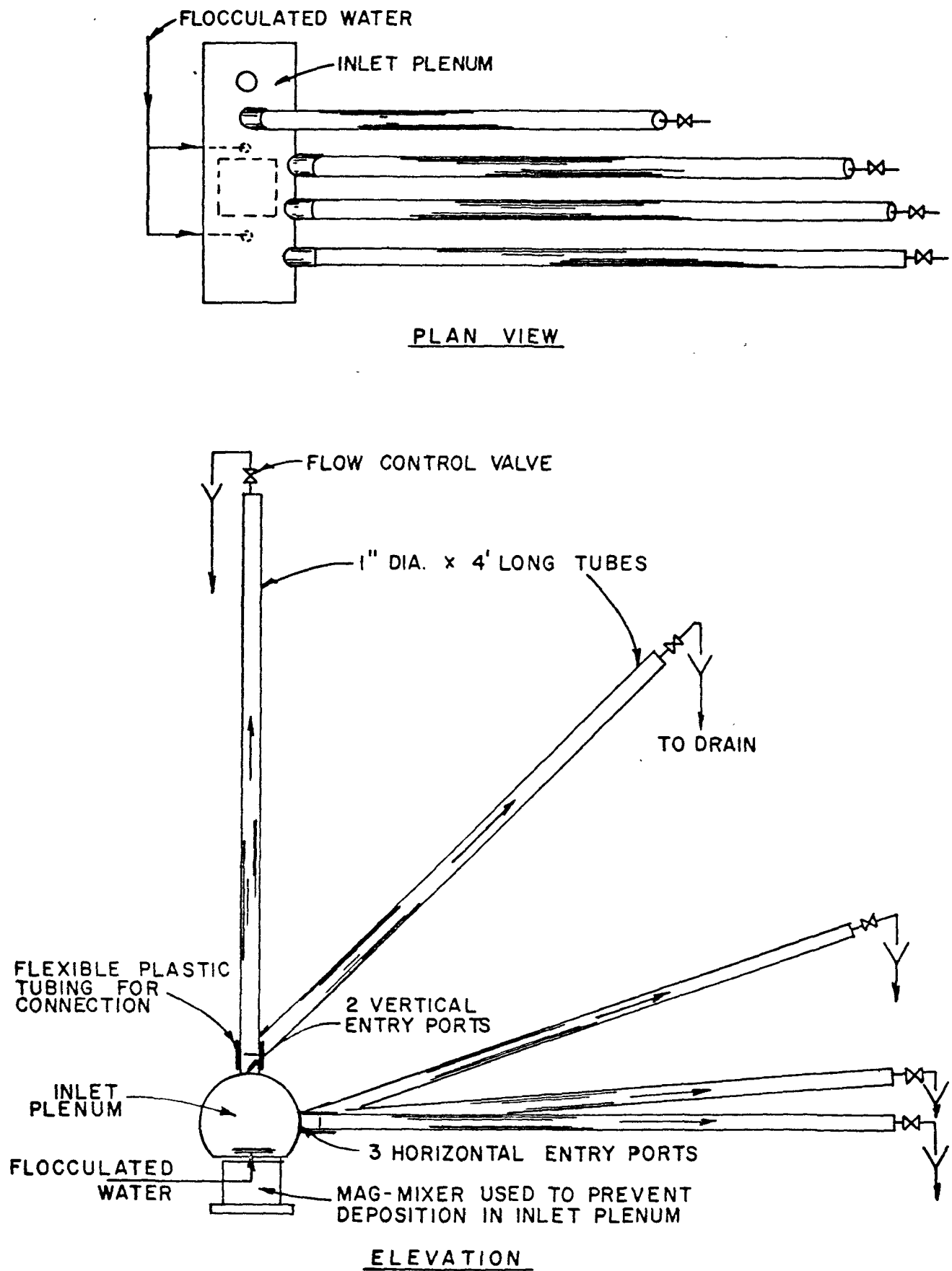
FIGURE 6

SCHEMATIC DIAGRAM OF TEST APPARATUS USED IN EVALUATING EFFECTS
OF TUBE INCLINATION
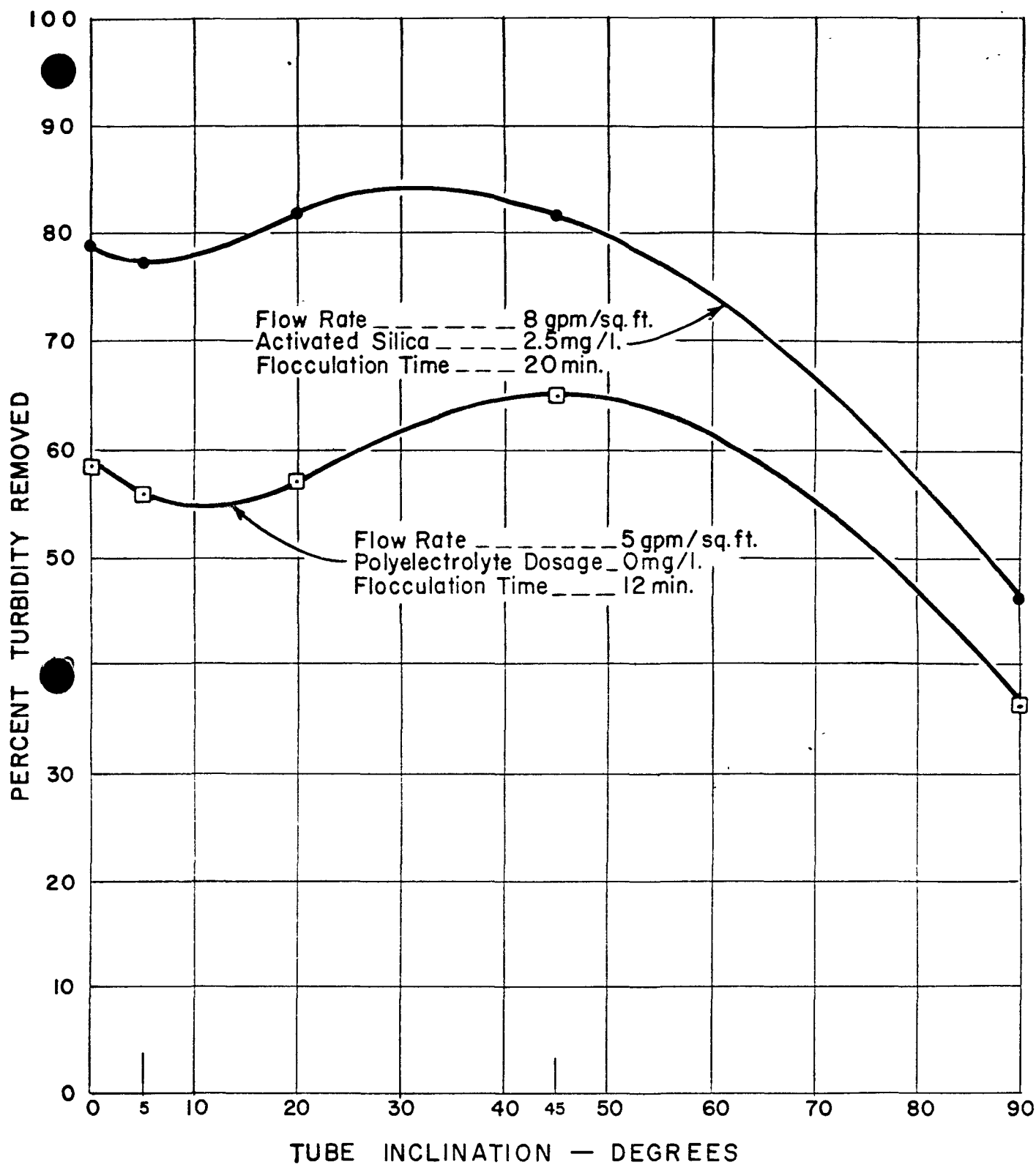
FIGURE 7

EFFECT OF TUBE INCLINATION ON SETTLING PERFORMANCE IN ONE INCH DIA., FOUR FOOT LONG TUBES

EXPERIMENTAL CONDITIONS

Tube Size _ _ _ _ _ _ _ _ 1" Dia., 4 feet Long
Ave. Influent Turbidity _ _ 150 J.U.

Flow Rate _ _ _ _ _ _ _ _ 2 gpm/sq.ft.
Polyelectrolyte Dosage _ _ 0 mg/l.
Flocculation Time _ _ _ _ 20 min.

□ Flow Rate _ _ _ 8 gpm/sq.ft.
Polyelectrolyte Dosage _ _ 0.5 mg/l.
Flocculation Time _ _ _ _ 20 min.

Flow Rate _ _ _ _ _ _ _ _ 5 gpm/sq.ft.
Polyelectrolyte Dosage _ _ 0 mg/l.
Flocculation Time _ _ _ _ 20 min.

PERCENT TURBIDITY REMOVED

100
90
80
70
60
50
40
30
20
10
0

35   40   45   50   55   60

TUBE INCLINATION — DEGREES

FIGURE 8

EFFECT OF TUBE INCLINATION ON SETTLER PERFORMANCE
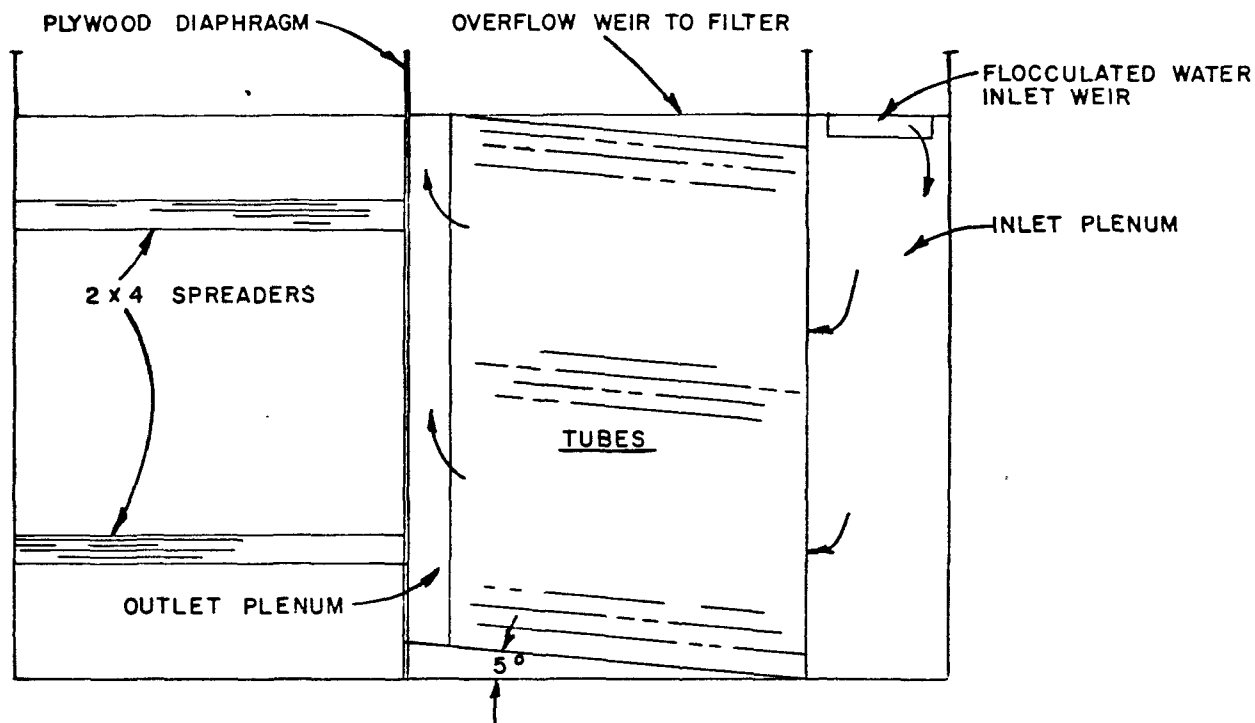
SECTION VIEW OF PLANT WITH TUBES INSTALLED IN UNIT AT
FIVE DEGREES
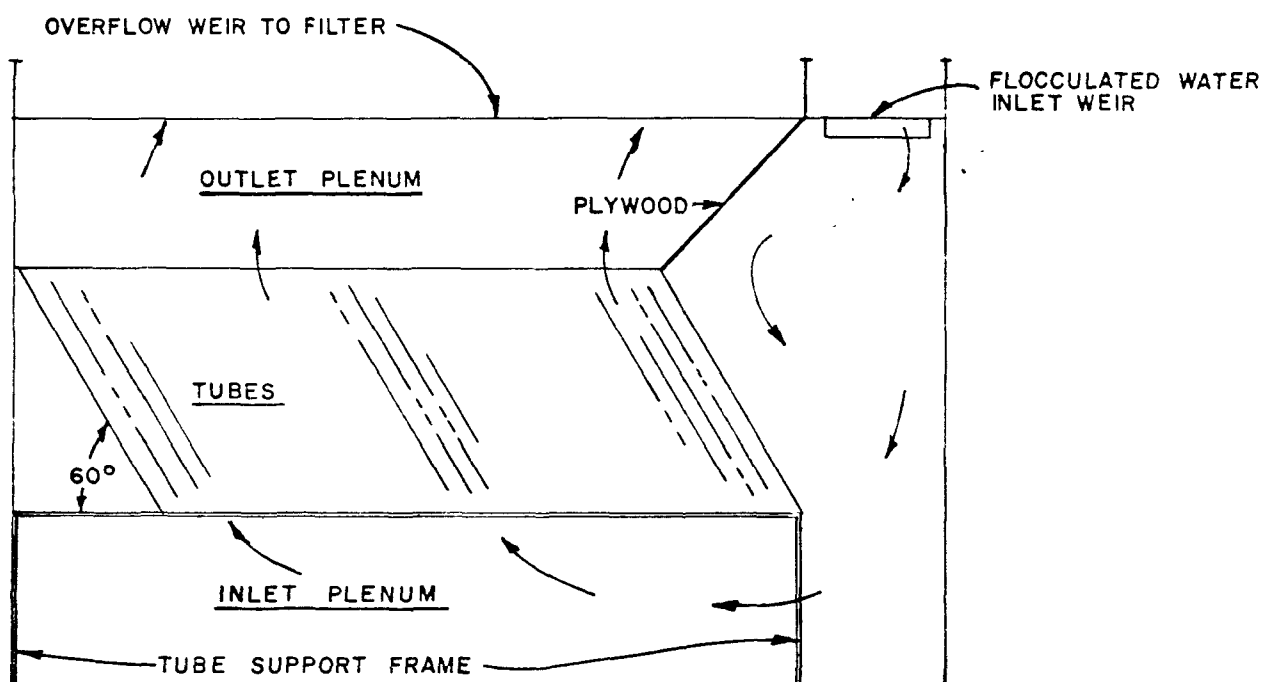
SECTION VIEW OF PLANT WITH TUBES INCLINED AT 60 DEGREES

FIGURE 9

APPARATUS USED IN PRELIMINARY FIELD EVALUATION OF
STEEPLY INCLINED TUBE SETTLER
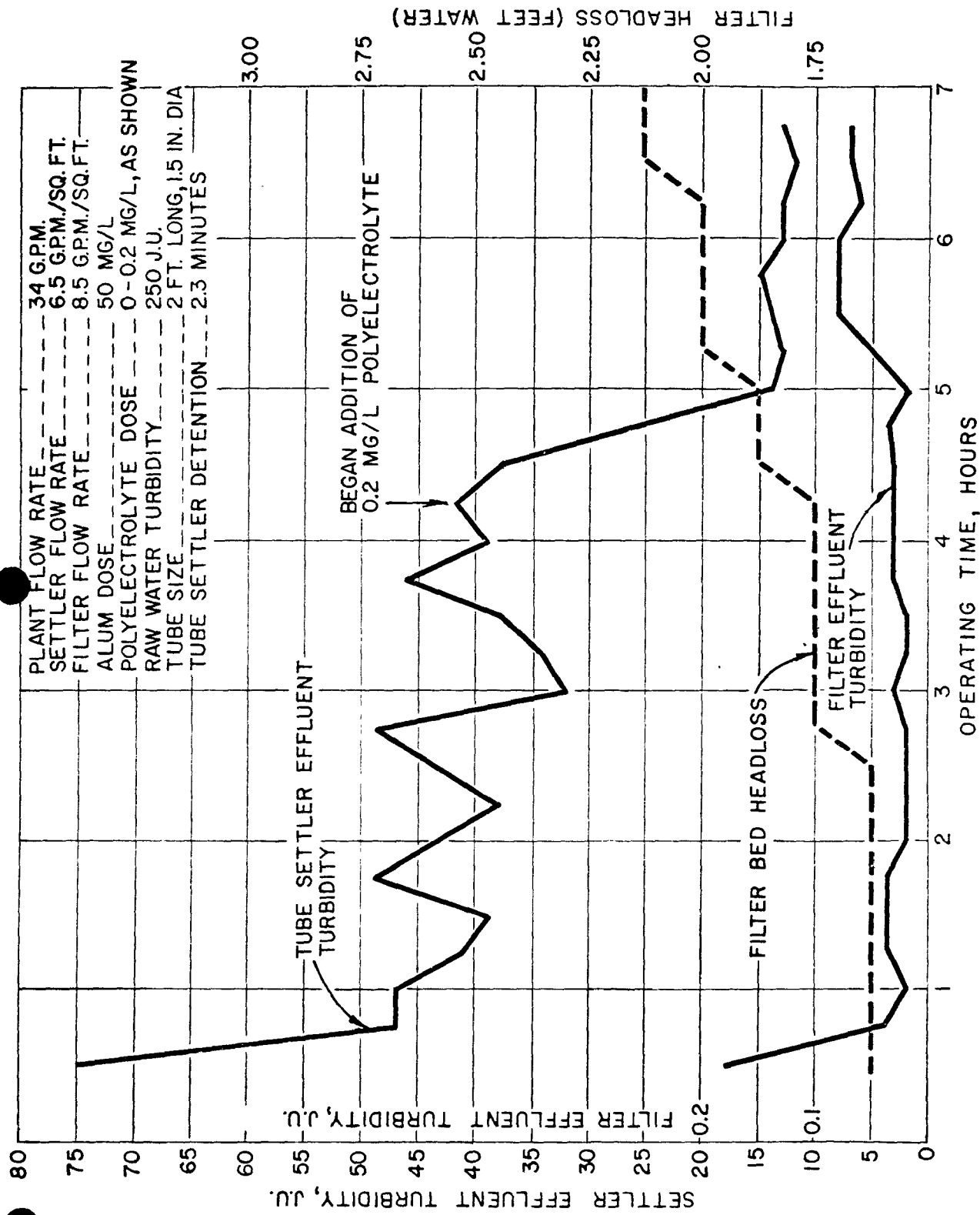
FIGURE 10

DATA FROM ONE RUN IN PRELIMINARY FIELD TESTS OF STEEPLY INCLINED
(60°) TUBES

FIGURE II

MODULE OF STEEPLY
INCLINED TUBES

FLOW MEASURING WEIR

WEIR BOX

POLYELEC-TROLYTE

ALUM

MUD

LET WATER

FLOCCULATOR DRIVE MOTOR

FLOCCULATOR TANK (280 GAL.)

AIR VENT

TO WASTE

EFFLUENT LAUNDER

SETTLING TUBES

SETTLING TANK (2'-0"x4'-6"x 6'-0")

SLUDGE DRAW OFF
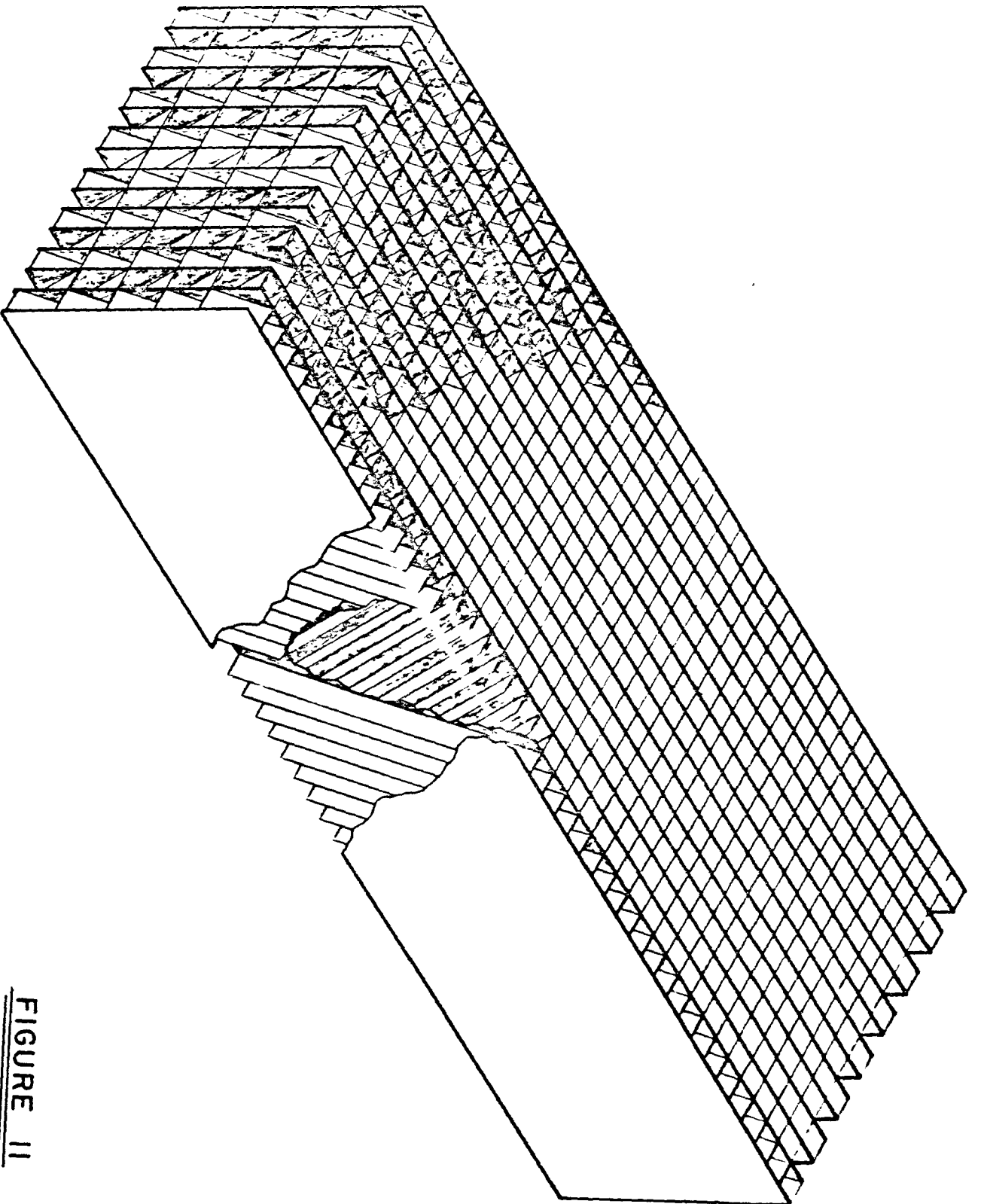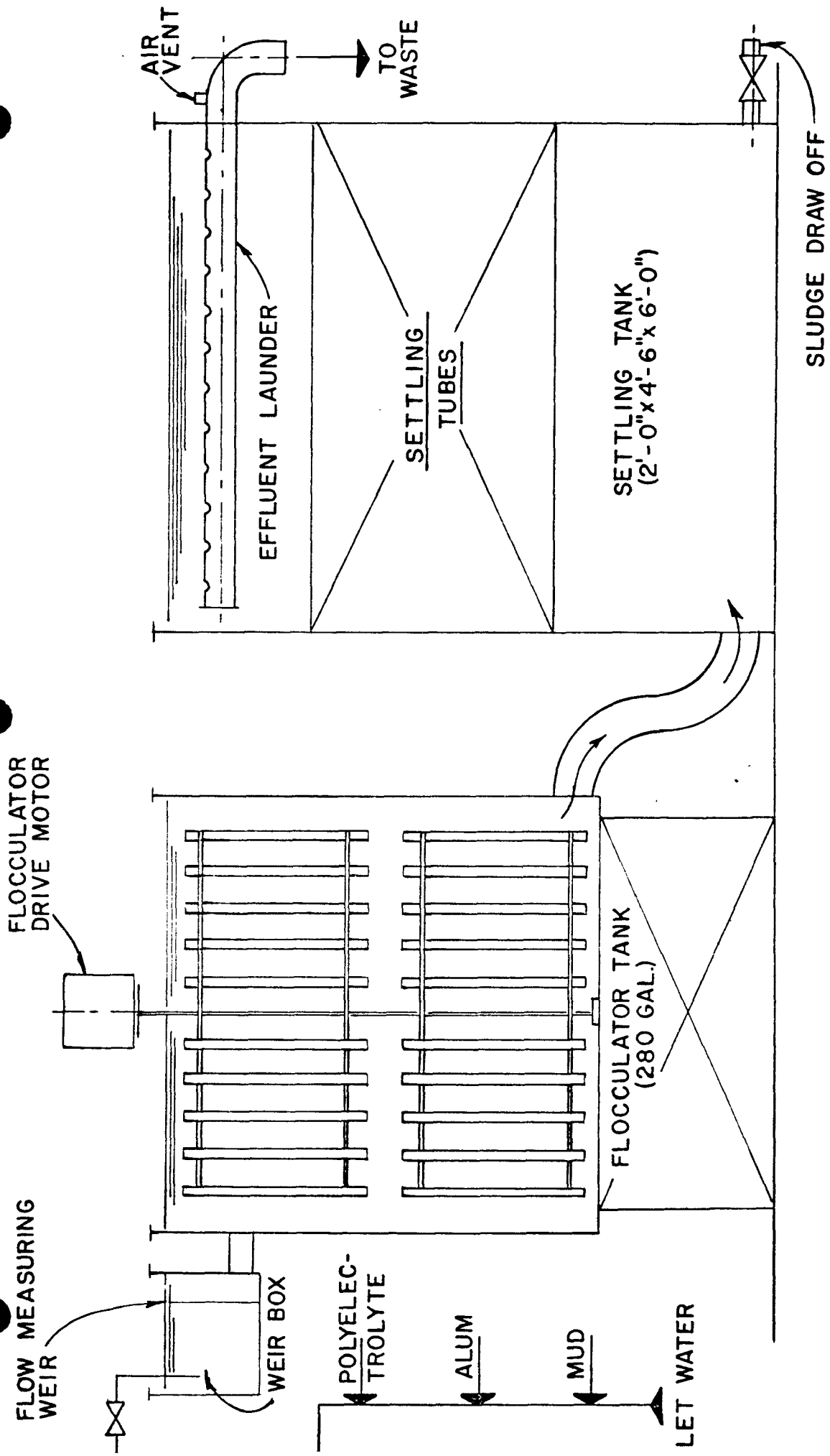
FIGURE 12

SCHEMATIC DIAGRAM OF FLOCCULATOR AND SETTLING TANK USED IN EVALUATING TUBE MODULE DESIGNS

FIGURE 13

NEWPORT, OREGON
WATER TREATMENT
PLANT
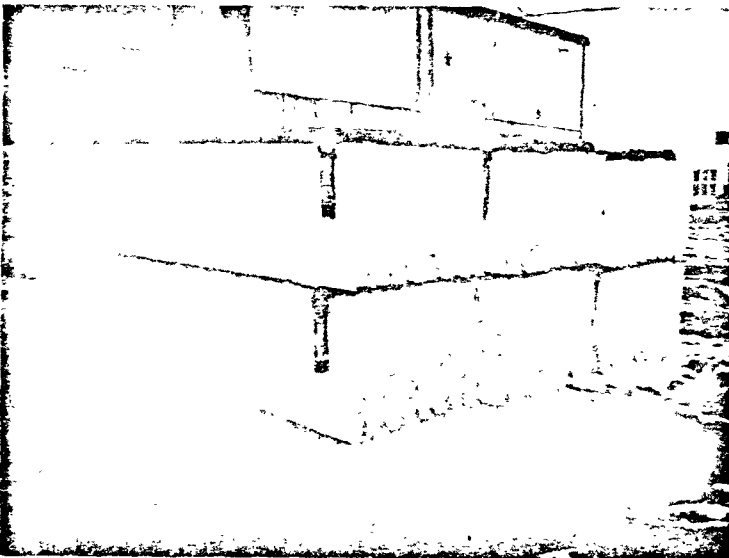
FIGURE 14

TUBE MODULES USED IN
NEWPORT CLARIFIER CONVERSION

BARRIER

SUPPORT MODULE

TUBE SETTLER
MODULES

30° SECTIONS
TYPICAL

A
B
C
D

A
B
C
D
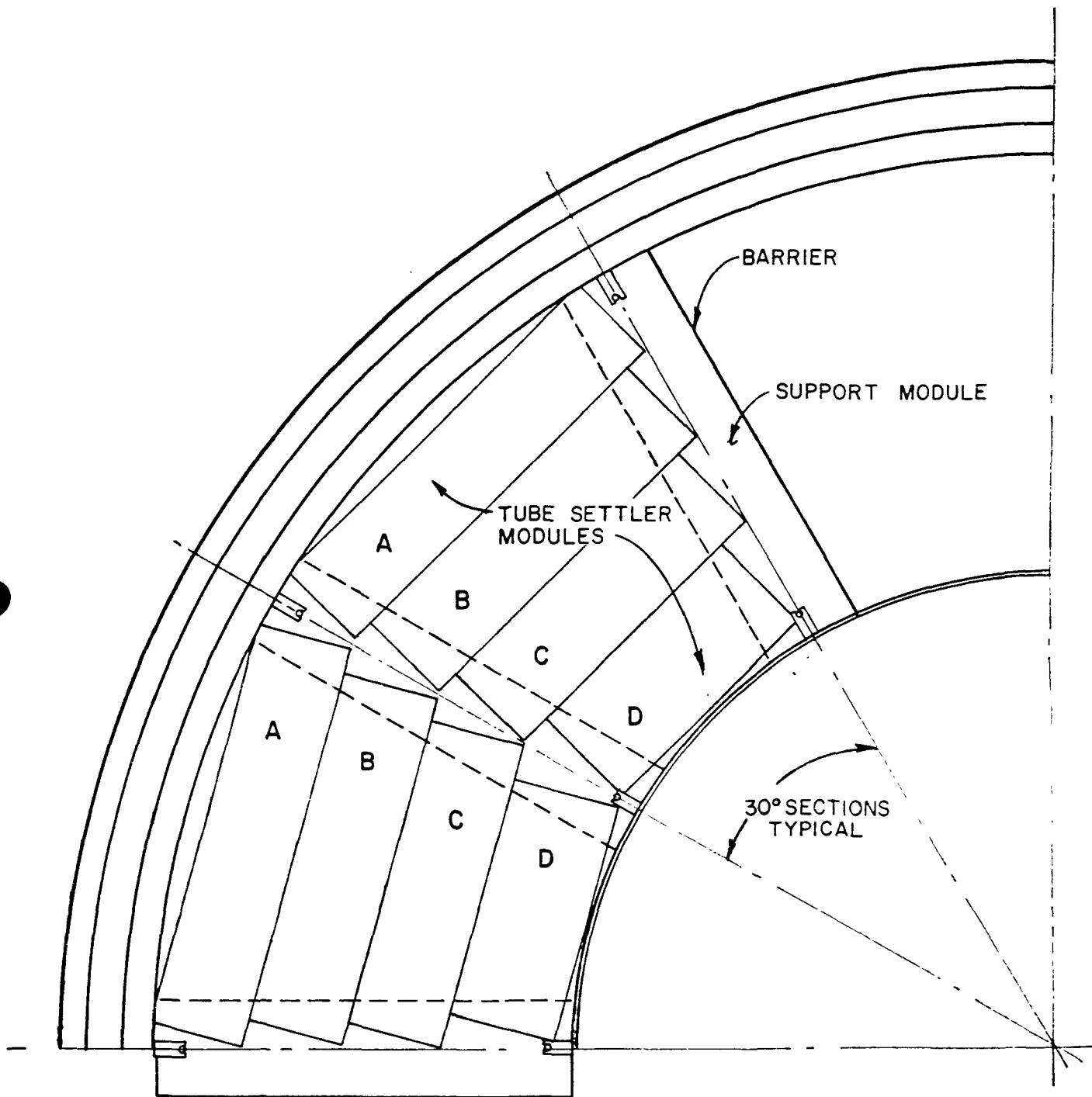
FIGURE 15

PLAN VIEW OF MODIFIED NEWPORT CLARIFIER

TUBE SETTLER
MODULES
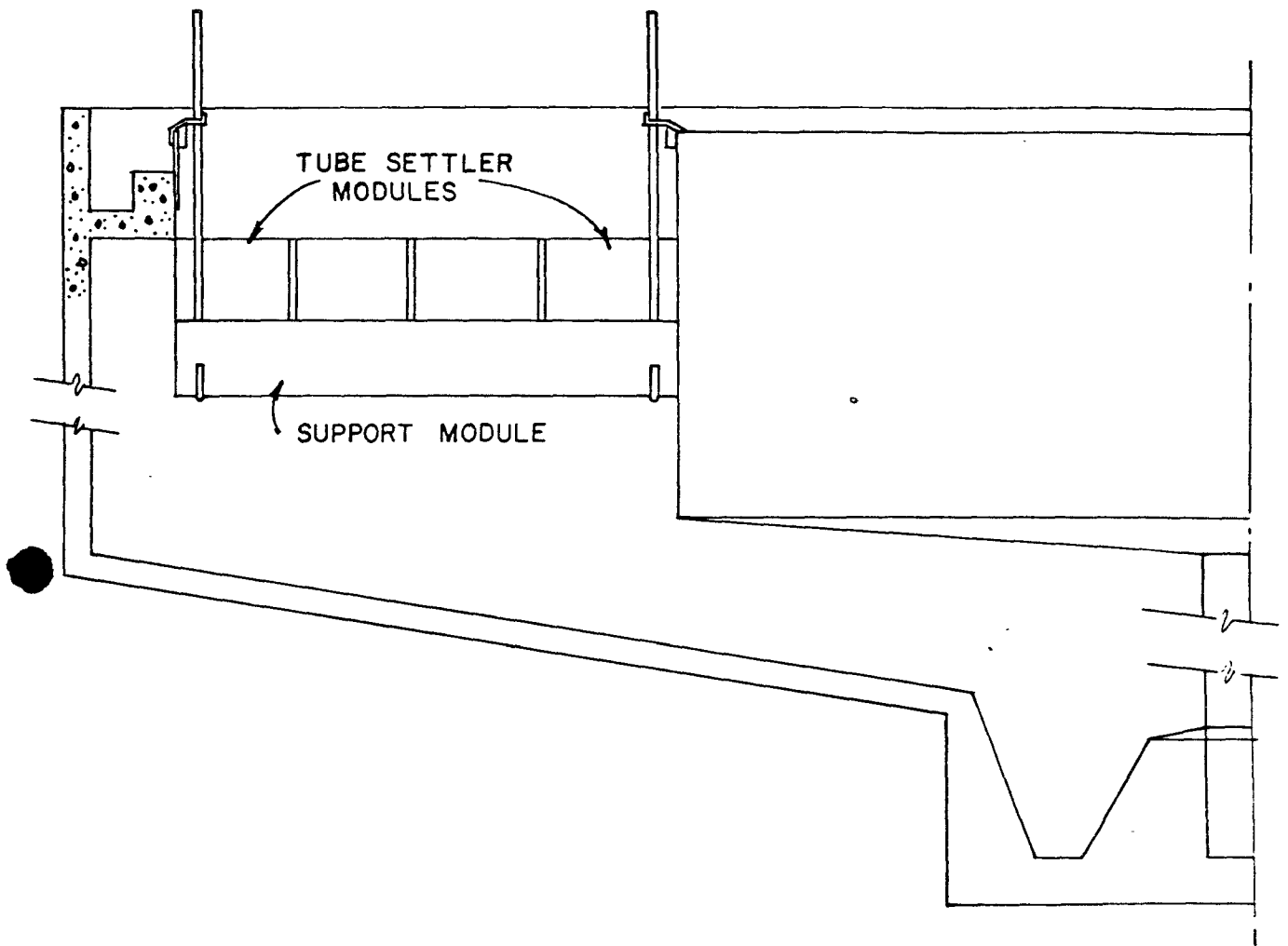
SUPPORT MODULE

FIGURE 16

SECTION OF MODIFIED NEWPORT CLARIFIER

FIGURE 17

SUPPORT MODULE
PRIOR TO INSTALLATION
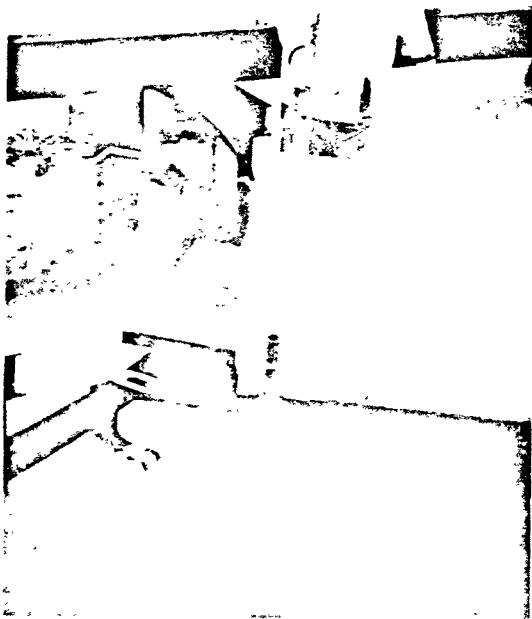
FIGURE 18

SUPPORT MODULE
BEING INSTALLED
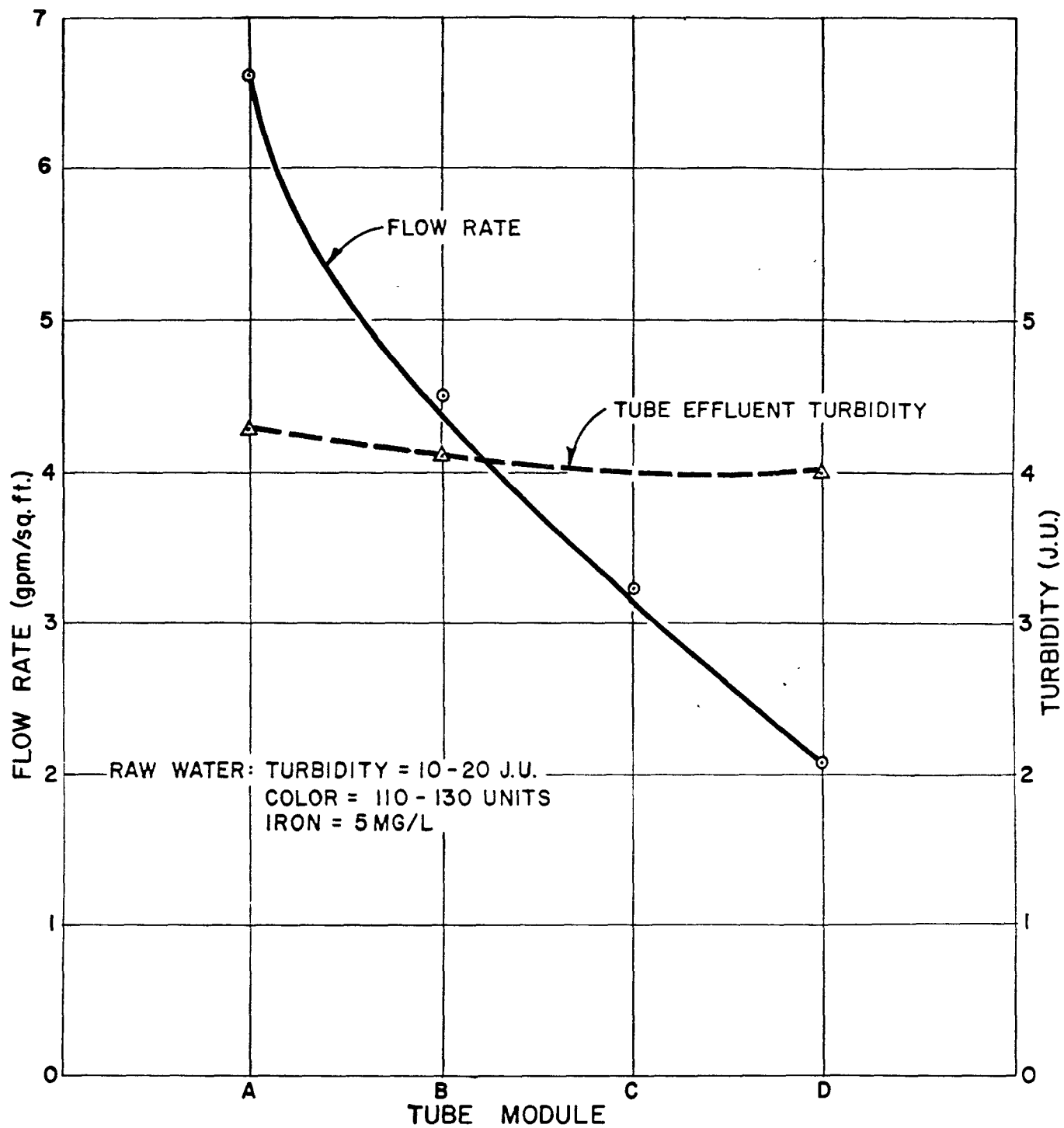
FIGURE 20

FLOW DISTRIBUTION AND TURBIDITY DATA IN INITIAL NEWPORT
TUBE MODULE INSTALLATION

HIGH RATE CLARIFICATION OF WASTE WATERS

Gordon L. Culp

Presented at the University of Kansas Sanitary Engineering Conference

January 3, 1968

The term "clarification" as used in this paper refers to removal of particulate matter from waste waters by either gravity sedimentation or filtration. The high-rate processes described in this paper enable reductions in size of sedimentation facilities by factors of 5-10 and allow filtration rates of 2-5 times those normally used.

FILTRATION

The desire to minimize pollution of the nation's water resources and the need to supplement these resources have made conventional secondary sewage treatment processes inadequate in some instances. Many of the tertiary treatment systems which have been proposed for removal of particulate and dissolved materials from secondary effluents involve filtration. Filtration of secondary effluent is a difficult problem in many respects. If the secondary effluent contains a high solids concentration, as many secondary effluents occasionally do, a conventional sand filter will blind at the surface in a very short time, even at low filtration rates. The reasons for this are apparent from an examination of Figure 1 which is a cross-section of the typical single media filter, such as a sand filter. During filter backwashing, the sand grades hydraulically with the finest particles rising to the top of the bed. As a result, most of the material removed by the filter is removed at or very near the surface of the bed. Materials passing the top few inches of the bed quite likely will pass completely through the filter. Only a small part of the total voids in the bed are used to store particulates and headloss increases very rapidly. When the secondary effluent contains relatively high solids concentrations, a sand filter will blind at the surface in only a few minutes.

One approach to increasing the effective filter depth is the use of a dual media bed using a discrete layer of coarse coal above a layer of fine sand, as shown in Figure 2. The work area is extended, although it still does not include the full depth of the bed, as there is some fine to coarse stratification within each of the layers, as shown by the graph depicting grain size. Effective size of the sand in a typical dual media filter is 0.4-0.5 mm.

It is apparent that an ideal filter would be the inverse of a sand filter; that is, it would have the coarsest material on top and the finest on the bottom, as illustrated by Figure 3. As shown on the right side of the

Figure 1
Cross-Section Through
Single-Media Bed
Such as Conventional
Rapid Sand Filter

Grain size
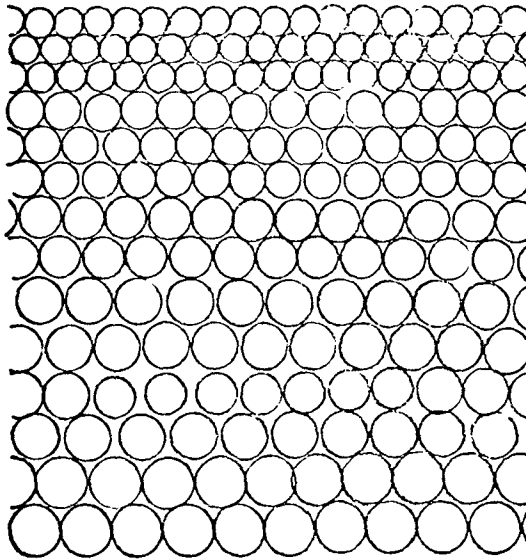
Figure 2
Cross-Section Through
Dual-Media Bed
Coarse Coal Above
Fine Sand

Grain size
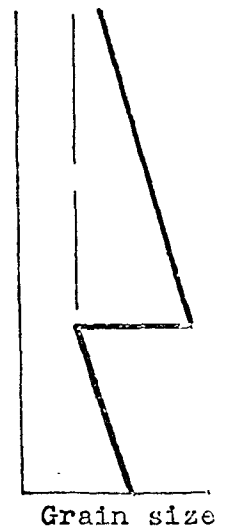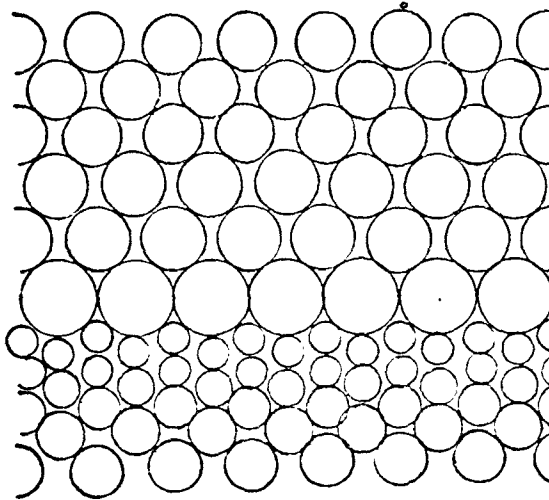
Figure 3
Cross-Section Through
Ideal Filter
Uniformly Graded From
Coarse to Fine
From Top to Bottom
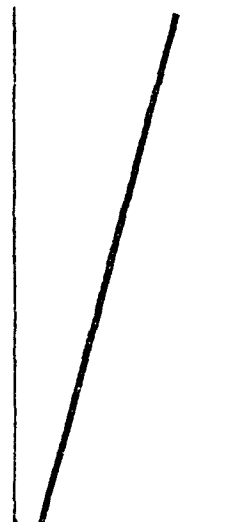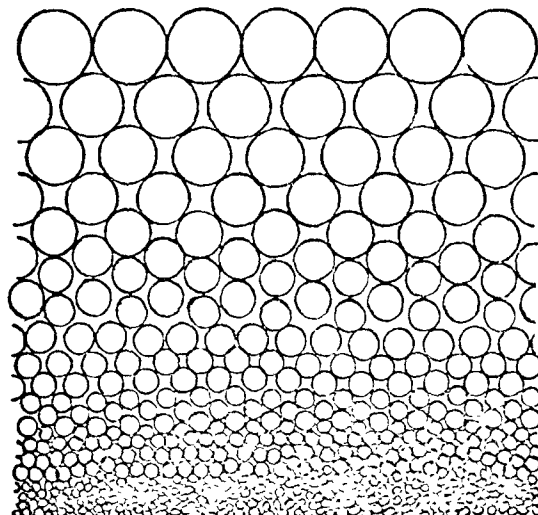
figure, the grain and pore sizes would be uniformly graded from coarse
to fine, from top to bottom. The author's company has developed a unique
method constructing a filter so that it closely approaches this ideal con-
figuration. In this "mixed-media" filter, three or more materials of dif-
fering specific gravities are allowed to intermix, and no attempt is made
to maintain separate or discrete layers of the different materials. The
three materials normally used are coal, sand, and garnet. Garnet has a
specific gravity of over 4 as compared to 2.6 for sand and 1.6 for coal.
The garnet, sand, and coal particles are sized so that some intermixing
of these materials occurs and no discrete interface exists between the
various materials. Of course, the differerences in specific gravity result
in the coarse, lighter coal occupying most of the upper portion of the
filter with the sand occupying most of the intermediate space between the
coal and bottom layer of very heavy garnet. The resulting filter has a
particle size gradation which decreases from about 1 mm at the top to about
0.15 mm at the bottom. The entire filter depth is utilized for floc removal
and storage. The very fine garnet used in the bottom layer of the filter
forces the effluent to pass through a much finer barrier than would be
provided by the coarser sand found in a coal-sand filter. The coarse upper
layer of the mixed-media filter greatly reduces its sensitivity to surface
blinding. Filter depths of 24-30 inches, filter flow rates of 5-10 gallons
per minute per square foot, and backwash rates of 15-20 gallons per minute
per square foot are normally used in secondary effluent filtration. Several
systems in which mixed-media filtration has been used for clarification
of secondary effluent are described in the next few paragraphs.

PLAIN FILTRATION

As shown in Figure 4, a very simple system for further clarification of
effluents is plain filtration. No chemical coagulation of the secondary
effluent is provided before it is passed through a mixed-media filter. The
efficiency of plain filtration is primarily dependent upon the degree of
biologic flocculation achieved in the secondary process. For example,
the biologic flocculation achieved in a trickling filter is relatively
poor and only 30-50 percent of the suspended solids found in a trickling
filter plant effluent will be removed by plain filtration. Conventional
activated sludge systems achieve a stronger biologic floc with the result
that 65-75 percent of the suspended solids in the secondary effluent can
be removed by plain filtration. The longer aeration periods used in
extended aeration plants achieve even a higher degree of biologic floccula-
tion, which enables more than 90 percent of the effluent suspended solids
to be removed by plain filtration. In an activated sludge system, our
research has shown that the efficiency of plain filtration is directly
proportional to the aeration time and inversely proportional to the
load factor (ratio of the amount of organic material added per day to the
amount of suspended solids present in the aeration chamber). Variation
of mixed liquor suspended solids in the normal operating range of 1,500-
5,000 milligrams per liter do not significantly effect the filterability
of the effluent at a given aeration time and load factor. With a load
factor of 0.15 and an aeration time of 12 hours, more than 90 percent of
the suspended solids in an activated sludge plant effluent have been removed
by plain mixed-media filtration at rates of 5-10 gallons per minute per
square foot.
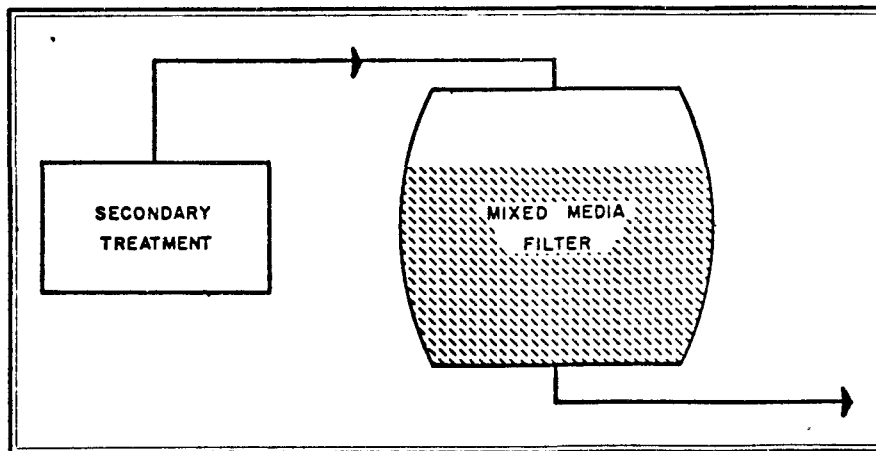
Figure 4

PLAIN FILTRATION
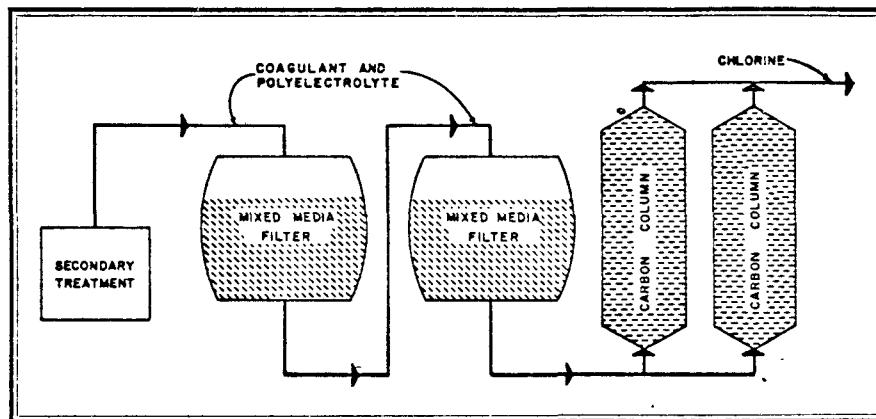
Figure 5

COAGULATION AND DIRECT FILTRATION
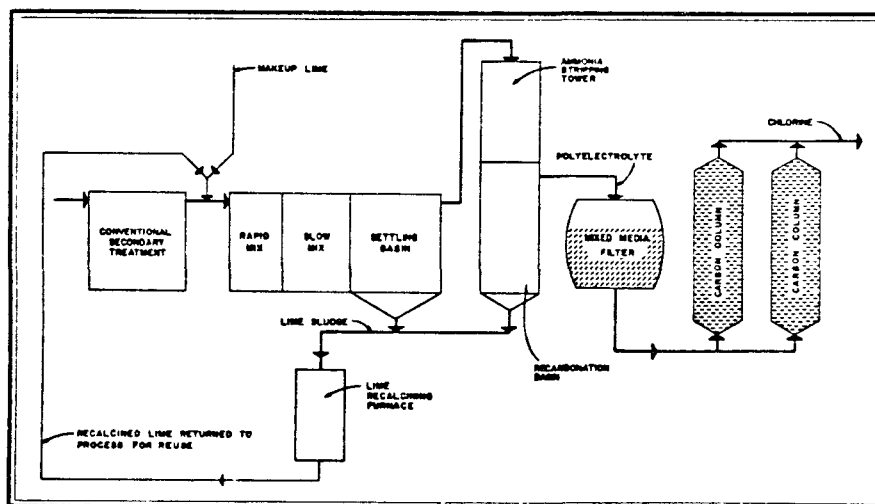
Figure 6

COAGULATION, SETTLING, AND FILTRATION

The anticipated effluent quality and the related cost of plain mixed-media filtration when applied to activated sludge effluent is shown in Table I. Table I also includes the anticipated effluent quality from tertiary systems providing higher degrees of treatment in a plant of this size and based upon a comparison, secondary treatment in a plant of this size and based upon the same assumptions, would be $80-$100 per million gallons. Of course, plain filtration alone removes only particulate matter and related BOD and has no significant effect on soluble organic materials or objectionable dissolved inorganic material such as phosphates. Dissolved organics may be reduced greatly by passage of the filter effluent through a column of granular activated carbon, as shown in Table I. The cost of activated carbon treatment is based on the assumption that the carbon is regenerated and reused which is essential to carbon treatment being economically feasible (1).

COAGULATION AND DIRECT FILTRATION

In order to provide an even further clarification of the secondary effluent and to provide reductions in the phosphate concentration, coagulation of the secondary effluent can be employed. With mixed-media filtration, it is possible to apply the coagulated secondary effluent directly to the filters. By using two filters in series as shown in Figure 5, it is possible to apply even the very high coagulant dosage required for phosphate removal directly to the filters. The process shown in Figure 5 is the original tertiary system used at the South Tahoe Water Reclamation Plant which has been described in detail elsewhere (1). In this case, the first filter consisted entirely of coarse coal while the second filter employed coal, sand, and garnet to provide coarse to fine gradation. The filter effluent is very low in suspended solids, BOD, turbidity, and phosphates, as shown in Table I. However, the filter effluent still contains soluble COD and color. Again, these soluble organics can be removed by passage of the filter effluent through columns of granular activated carbon.

If the secondary effluent quality is subject to wide fluctuations, the reliability of the tertiary system will suffer when direct filtration is employed. Even the mixed-media filters cannot tolerate suspended solids concentrations of several hundred mg/l which may occur during a severe upset of the secondary plant. To increase the reliability of the system under these conditions, the use of supplemental settling between the secondary system and the tertiary system is desirable.

COAGULATION FOLLOWED BY SETTLING AND FILTRATION

A schematic diagram of the process now being used at the South Lake Tahoe plant, which has recently been expanded from 2-1/2 MGD to 7-1/2 MGD, is shown in Figure 6. The secondary effluent is chemically coagulated, flocculated, and then passed through a settling basin, prior to filtration and carbon treatment. Thus, if the secondary system should upset severely, most of the biological solids will be trapped in the intermediate settling basin and will not adversely effect the tertiary system. The coagulant shown in Figure 6 is lime rather than alum which was used in the original Tahoe

TABLE I

COST OF TERTIARY PROCESSES TO PRODUCE VARIOUS QUALITIES
OF RECLAIMED WATER FROM SECONDARY EFFLUENT

	TOTAL (1) COST FOR 10 MGD PLANT	TYPICAL QUALITY						
		SUSPENDED SOLIDS (mg/1)	BOD (mg/1)	COD (mg/1)	TURBIDITY (mg/1)	COLOR	PO_4 (mg/1)	TOTAL N(mg/1)
Plained Mixed-Media Filtration	$ 10/mg	1-5	1-5	40-70	0.3-5	60-80	20-40	20-40
plus Activated Carbon	$ 50/mg	<1-3	<1	5-15	0.3-3	<5	20-40	15-35
Coagulation plus Direct Mixed-Media Filtration (2)	$ 80-$100/mg	<1	<1	30-60	0.1-1.0	10-30	0.1-1.0	15-35
Coagulation plus Sedimentation	$ 60-$ 80/mg	1-5	1-3	40-70	1-2	10-30	1-2	15-35
plus Filtration	$100-$120/mg	<1	<1	30-60	0.1-1.0	10-30	0.1-1.0	15-35
plus Activated Carbon	$135-$155/mg	<1	<1	1-15	0.1-1.0	<5	0.1-1.0	10-30
plus Ammonia Stripping	$150-$170/mg	<1	<1	1-15	0.1-1.0	<5	0.1-1.0	1-2

(1) Cost includes amortization of capital cost over 20 years at 4 percent interest, operation, maintenance costs.

(2) Costs shown are for coagulant doses required for phosphate removal (200 mg/1 alum).

tertiary plant. The change in coagulant resulted from extensive studies
on the feasibility of recovery and reuse of both alum and lime. No economically
feasible method of alum recovery and reuse could be found, while it was
found that lime could be recalcined and reused (1). The high lime doses
required for phosphate removal raise the pH of the secondary effluent to
such high levels (greater than 10.5) that the ammonia nitrogen in the lime
coagulated effluent may be removed as ammonia gas. This can be accomplished
by passing the high pH effluent through a packed tower in which the ammonia
gas is liberated and stripped from the system by a countercurrent flow of
air. Details of the studies on ammonia stripping have been published elsewhere (1
The stripping tower effluent is recarbonated and then passed through the
filtration and activated carbon systems. As can be seen from Table I,
this system provides an effluent with very low concentrations of suspended
solids, BOD, COD, turbidity, color, phosphate, and nitrogen.

SEDIMENTATION

The removal of particulate matter by gravity sedimentation is one of the
most widely used processes in the field of waste treatment. Investment
in settling tanks represents a significant portion of the total capital
costs involved in treatment facilities. In spite of the importance of this
process, application of the principles involved has progressed so slightly
that design criteria established more than 50 years ago are still in wide
spread use. Settling basin design, for the most part, continues to conform
to previous practice in size and shape with little consideration for the
principles involved. As indicated in the above discussion, gravity settling
is not only important in primary and secondary systems but can play an
important roll in tertiary treatment systems.

Our research program has devoted a great deal of time to develop a method
by which the basic theories of sedimentation can be applied in a practical
manner. This basic theory indicates that the use of very shallow settling
basins would enable the detention time of the settling process to be reduced
to only a few minutes, in contrast to the several hours used in conventional
basins. The basic theory can be readily understood by examination of
Figure 7. This figure is a schematic representation of the cross-section
of an ideal, rectangular, horizontal flow settling basin of depth, h_o,
and length, L_o, with the direction of flow being from left to right. The
diagonal dashed line represents the settling path which a particle with
horizontal velocity V and settling velocity v_o would take, if it enters at
the top of the basin. This particle would strike the bottom of the basin
with dimensions h_o and L_o and would be removed. A particle with a slower
settling velocity, such as v_s, would not be removed, but could be if a
false bottom or tray were inserted at depth h. It is apparent that as
basin depth h is reduced even further that basin length L could also be
reduced. This basic theory has been accepted for more than 50 years (2)
and indicates that settling efficiency is independent of basin detention
time and is related only to basin overflow rate and basin depth. In fact,
in 1904, Hazen (2) pointed out that detention times of 10 minutes would
be more than adequate if tray spacings as low as one inch could be provided.
Numerous attempts to apply this theory in the water and waste treatment
field have been made in the last 50 years. The problems which limited
the success of these attempts were sludge removal and flow distribution.
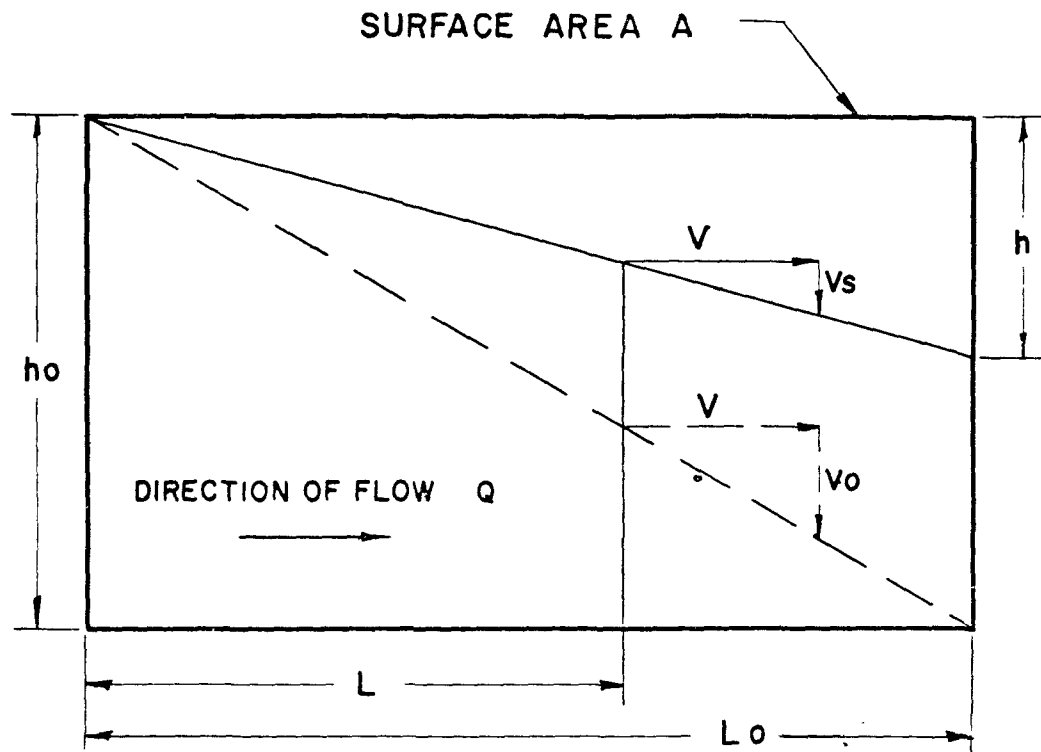
FIGURE 7

IDEALIZED SETTLING PATHS OF DISCRETE PARTICLES IN A
HORIZONTAL FLOW TANK

The minimum tray spacing was limited by the space required for insertion
of mechanical sludge removal devices. The wide shallow trays caused flow
distribution problems by their inherently unstable hydraulics. The sub-
stitution of relatively small diameter tubes for the shallow wide trays
overcomes these problems.

Longitudinal flow through tubes with a diameter of a few inches offers
theoretically optimum hydraulic conditions for sedimentation. Such tubes
have a large wetted perimeter relative to the wetted area and thereby provide
laminar flow conditions as evidenced by very low Reynolds numbers.

For example, a one-inch diameter tube, 4 feet long, through which water is
passed at a rate of 10 gpm per square foot of cross-sectional area would
have a Reynolds number of 24, an equivalent overflow rate of 235 gpd per
square foot, and a detention time of only 3 minutes.

Perhaps the manner in which the tube settling device functions can be most
easily understood is by studying Figure 8, in which four demonstration tube
settlers are pictured. The particular tubes shown in Figure 8 are
1-inch diameter, 2 foot long, plexiglass tubes through which flow is passing
from left to right. The influent consists of an alum coagulated raw water
with an initial turbidity of 300 Jackson Units. The total detention time
in the tube settling apparatus is 3 minutes with a velocity of 0.011 feet
per second in each tube. The upper tube in the photograph has just been
placed in service. The sludge deposit near the entrance to this tube has
built up to the point where the velocity in the small inner space between
the top of the sludge deposit and top of the tube is too great to permit
a greater depth of sludge deposit. The second tube from the top has been
operating for a longer period and the sludge deposits have filled a greater
portion of this tube. In these particular tubes, the floc must fall an
inch or less to be removed from the system and, of course, this requires
much less time than it would for the floc to fall the several feet required
in a conventional settling basin. The bottom two tubes have been running
for longer time intervals and it can be seen that essentially the entire
tube volume is used for floc removal and floc storage before any significant
carryover of floc begins to occur.

The method of sludge removal is very simple and eliminates the requirement
for mechanical sludge removal equipment. The tubes are inclined upwardly
slightly in the direction of flow. Periodic draining of the tube reservoir
has been found to be adequate for sludge removal. As the reservoir is
drained, the water level falls at such a rate that it hydraulically scours
the sludge deposits from the tubes. These slightly inclined tubes are
generally operated in conjunction with a filter so that the volume of water
drained from the tubes during the sludge removal process may be replaced
by a portion of the filter backwash water. A detailed discussion of the
tube settling process can be found elsewhere (3, 4). The method of operation
of the tube settling system should become clear as we examine some waste
treatment systems in which it has been integrated.
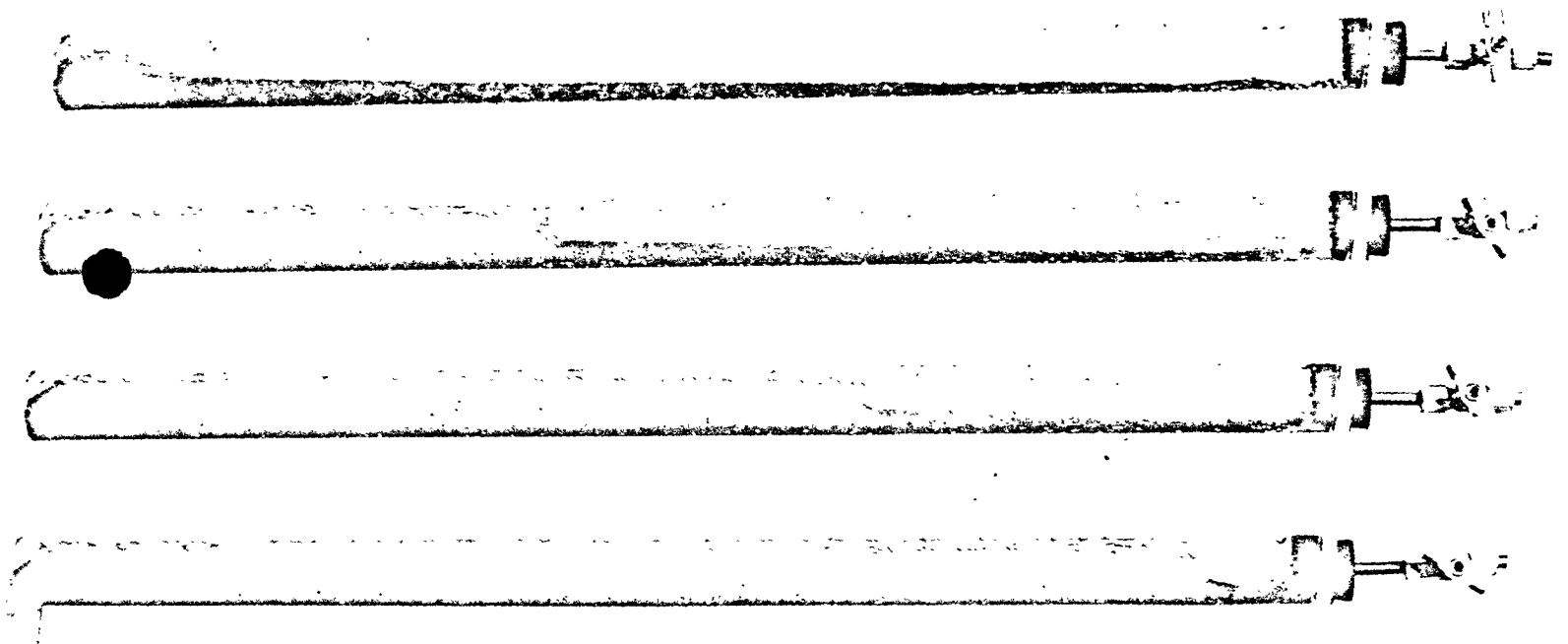
Figure 8

SLUDGE DEPOSITION IN FOUR DEMONSTRATION TUBE SETTLERS

EFFLUENT POLISHING

Figure 9 shows an effluent polishing process in which the tube settling principle has been utilized. The particular system shown provides polishing of the effluent from a package sewage treatment plant. These small package plants frequently suffer from periodic discharge of high suspended solids concentrations in the plant effluent. These plant upsets may result from inadequate sludge wasting by the plant operator, surges in hydraulic flow through the plant, or failure of the plant sludge collection and return system. The package plant effluent is collected in a sump and then pumped through a tube settling unit followed by mixed-media filtration. The filter effluent is collected in a storage tank and is used for backwashing the filter. The backwash storage tank also serves as a chlorine contact basin. The role of the tube settler in the process shown in Figure 9 is that of a supplemental solids separation device which allows the filter to continue to operate efficiently even during severe upsets of the package sewage treatment plant. During such upsets, any attempt at direct filtration of the package plant effluent would meet with failure due to rapid blinding of the filter surface with the extremely high solids concentrations encountered. During such an upset, using the system shown in Figure 9, the bulk of the solids contained in the package plant effluent will be removed in the tube settler and the clarified tube settler effluent will then be amenable to mixed-media filtration. When the sludge storage capacity of the tube settler is exhausted, a significant amount of floc carryover will rapidly occur and cause the headloss on the filter to increase to a point such that the backwashing cycle is automatically initiated by filter headloss. During the backwash cycle, filter effluent is withdrawn from a backwash storage tank, passed up through the mixed-media filter, and recycled to the aeration tank. At the same time, the drain valve on the tube settler is opened and the entire contents of the tube basin drain into the sludge holding tank. The solids captured in the tube settler are completely scoured from the tubes during this draining cycle. Prior to the end of filter back-washing, the last portion of filter backwash water is diverted into the tube reservoir to replace the water drained from the basin. As soon as the 7 minute cleaning cycle is completed, the polishing unit is immediately returned to service. During the cleaning cycle, the incoming sewage flow is stored in the collection sump along with the backwash water recycled to the system.

Detailed data on such a polishing unit have been published (5) and show that such a system will produce average effluent suspended solids and BOD of less than 5 mg/l. The polishing process continues to function efficiently even though the solids concentrations in the package plant effluent may be in excess of 2,000 mg/l. Operational data from a 20,000 gpd system which has been in operation for over one year at a U.S. Forest Service installation in Oregon (6) show that the periodic heavy solids losses normally encountered in a package plant installation can be completely eliminated by use of this polishing process. Data from this Forest Service installation have also shown that in addition to providing very low BOD and suspended solids concentrations, the low turbidity of the polishing process effluent allows very effective disinfection. Coliform concentrations of less than 2 per 100 ml are routinely achieved. Of course, the extremely short residence times in the tube settlers make the polishing process a very compact unit which occupies a very small portion of land compared to the secondary process.
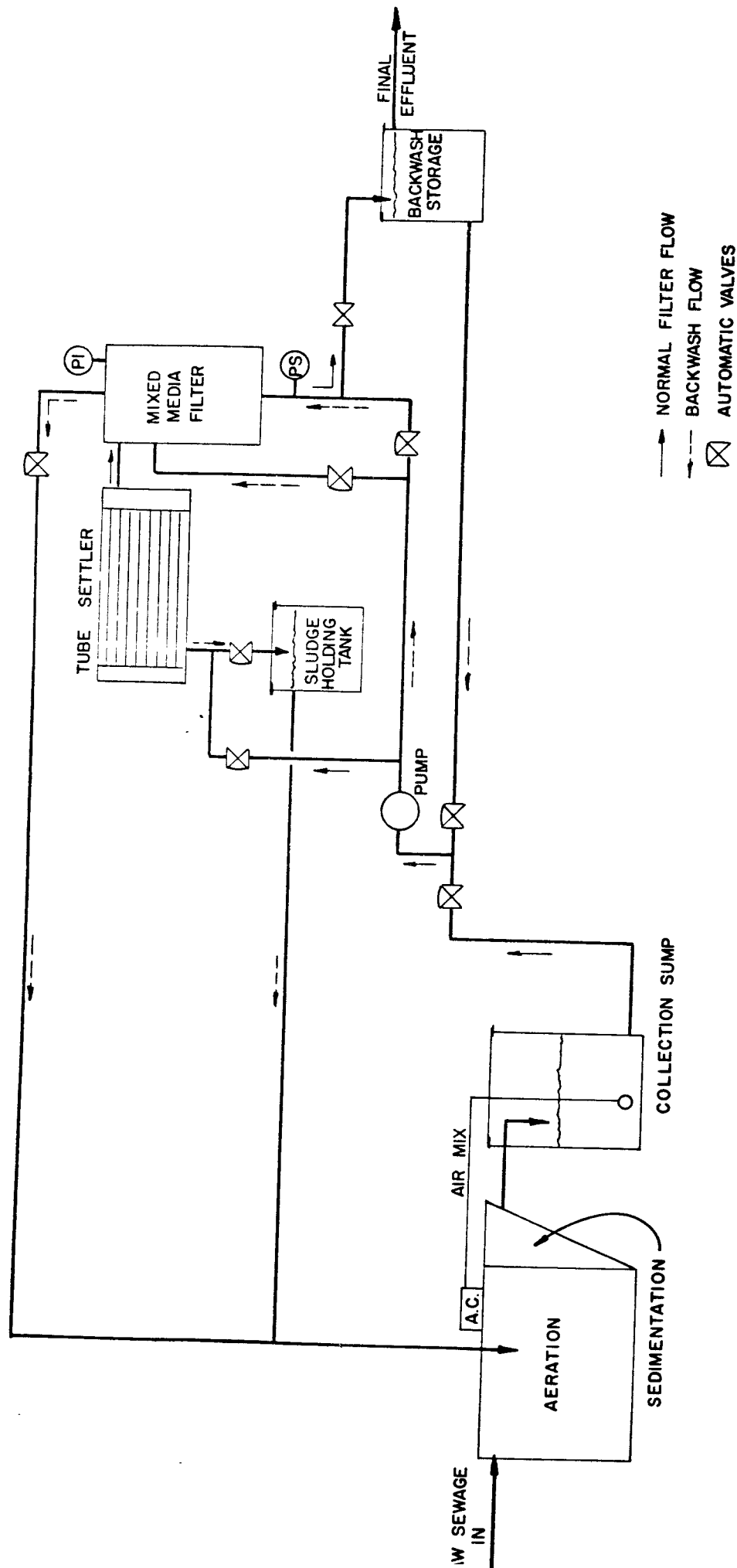
FIGURE 9

SCHEMATIC DIAGRAM OF EFFLUENT
POLISHING PROCESS EMPLOYING
TUBE SETTLING

Phosphate removal can also be achieved in a packaged polishing system using the tube settling, mixed-media filter combination. Such a system is shown schematically in Figure 10. In this system, the effluent receives alum coagulation and flocculation prior to entering the tube settler. As in the polishing process described above, the very low residence times, (less than 15 minutes) in the tube settling reservoir allow very compact tertiary systems to be manufactured. This is well illustrated by the plant pictured in Figure 11. This is a 100 gpm plant providing flocculation, sedimentation, and filtration with a total plant residence time of less than 30 minutes. The overall plant dimensions are 6 feet by 6 feet by 14 feet. A 30,000 gpd plant similar to that pictured in Figure 11 has been in operation for several months at a Federal Water Pollution Control Administration project in Minnesota. Effluent quality shown in Table 1 for coagulated and filtered secondary effluent is readily achieved in this package system.

CLARIFICATION OF MIXED LIQUOR

The tube settling systems described thus far have all employed tubes inclined only slightly upwardly in the direction of flow so that they may be cleaned by periodic drainage. As has been described above, the tubes are generally used in conjunction with a filter so that the volume of water drained during the sludge cleaning cycle can be replaced by a portion of backwash water. During tests of the effects of angle of tube inclination it was found that if the tubes were inclined steeply enough, the solids which settle to the bottom of the tube will slide downward along the tube bottom and eventually exit the tube. It has been found that at angles in excess of 45 degrees, a natural evacuation of the solids can be achieved continually without interupting the flow through the tubes. A countercurrent flow pattern is established in which particles entering the tube are carried upwardly until they settle to the lower tube surface at which point they become trapped in a downward flowing stream of concentrated solids. The advantages of such a self-cleaning high-rate sedimentation device are apparent.

One application of interest is the separation of the high mixed liquor solids levels encountered in activated sludge systems. Figure 12 is a schematic diagram of a pilot plant used to evaluate the steeply inclined tubes for continuous separation of mixed liquor. Comminuted raw sewage was aerated with the resulting mixed liquor passing through a tube settling basin with a total detention time of less than 15 minutes. The tube effluent was then passed through a mixed-media filter for further polishing. It was found that the tube settler provides very efficient separation of the mixed liquor solids as well as a continuous gravity return of the solids to the aeration basin. Mixed liquor solids in excess of 7,000 mg/l have been maintained with the system shown in Figure 12. The tube settler effluent solids have averaged about 40 mg/l. Filtration of the tube settler effluent consistently reduces the solids and BOD to less than 5 mg/l.

A prototype of a package system incorporating the essential features of Figure 12 has been in operation for several months. An aeration time of 12 hours, a tube settling detention time of 15 minutes, and an average filtration rate of 5 gpm per square foot have been used. A reduction in aeration time over that used in conventional extended aeration package plants allows the very high quality effluent to be produced from an integrated package of smaller size than a conventional package plant of the same capacity. For example, a 20,000 gpd extended aeration plant would occupy 42 feet
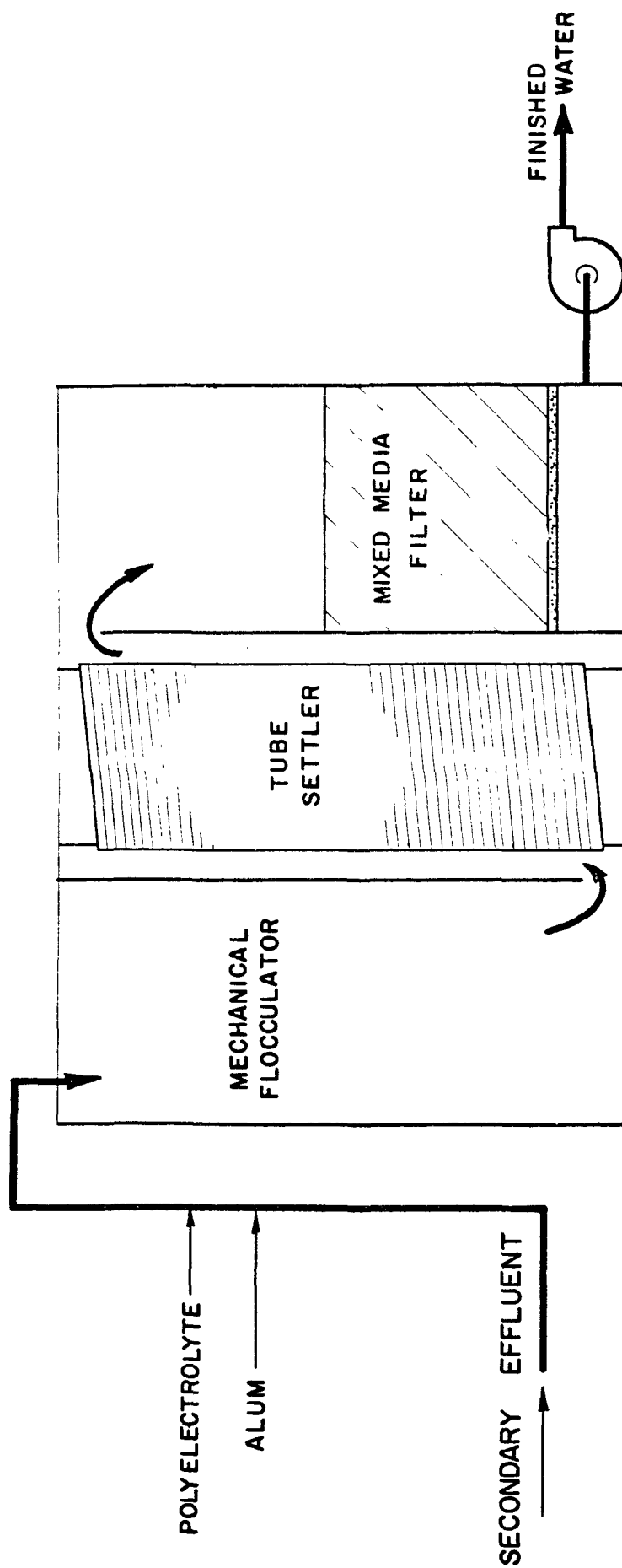
FIGURE 10

SCHEMATIC DIAGRAM OF PACKAGE TERTIARY TREATMENT PLANT

-15-
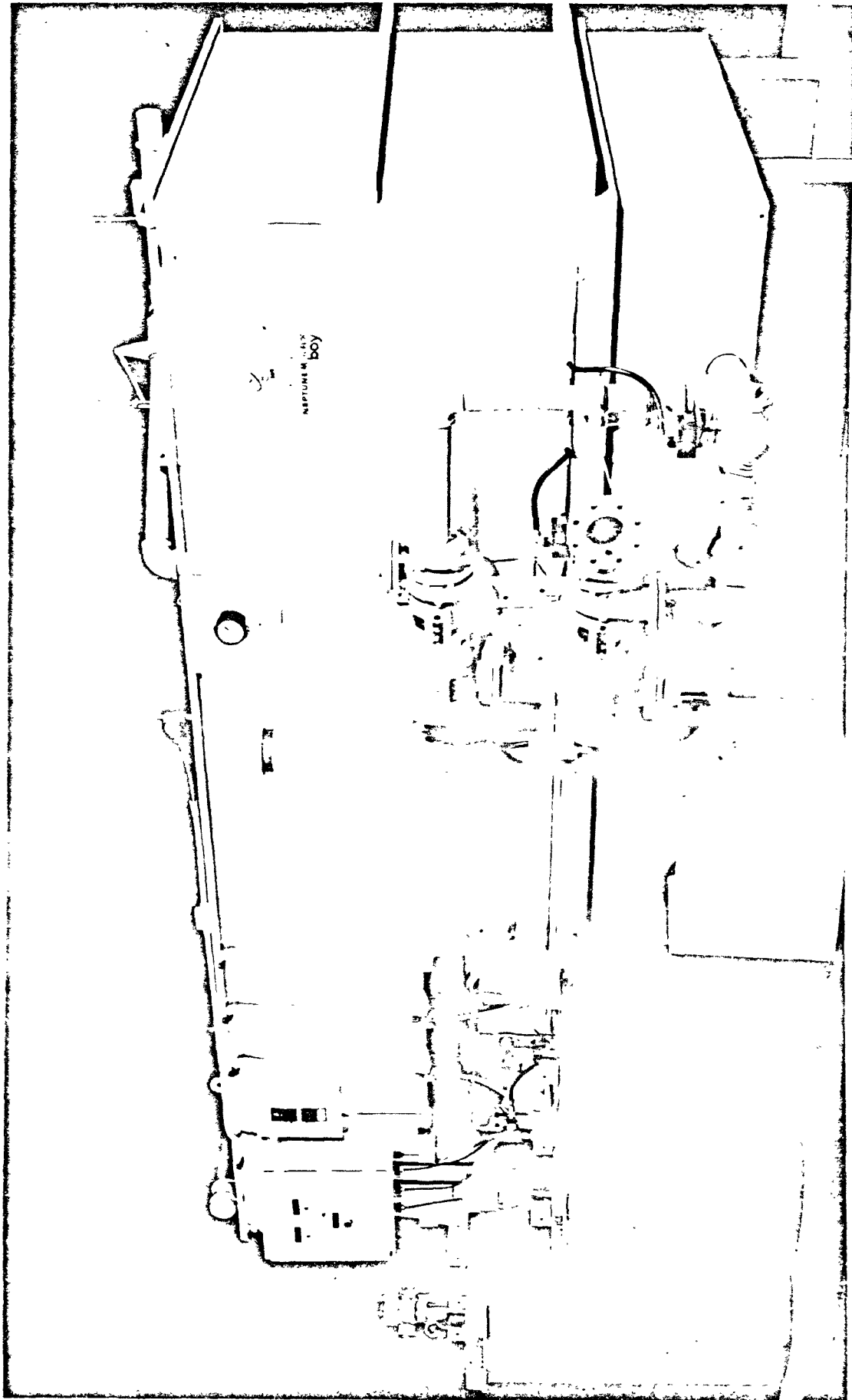
Figure 11

100 GPM PACKAGE TERTIARY TREATMENT PLANT
INCLUDING FLOCCULATION, SETTLING, AND FILTRATION
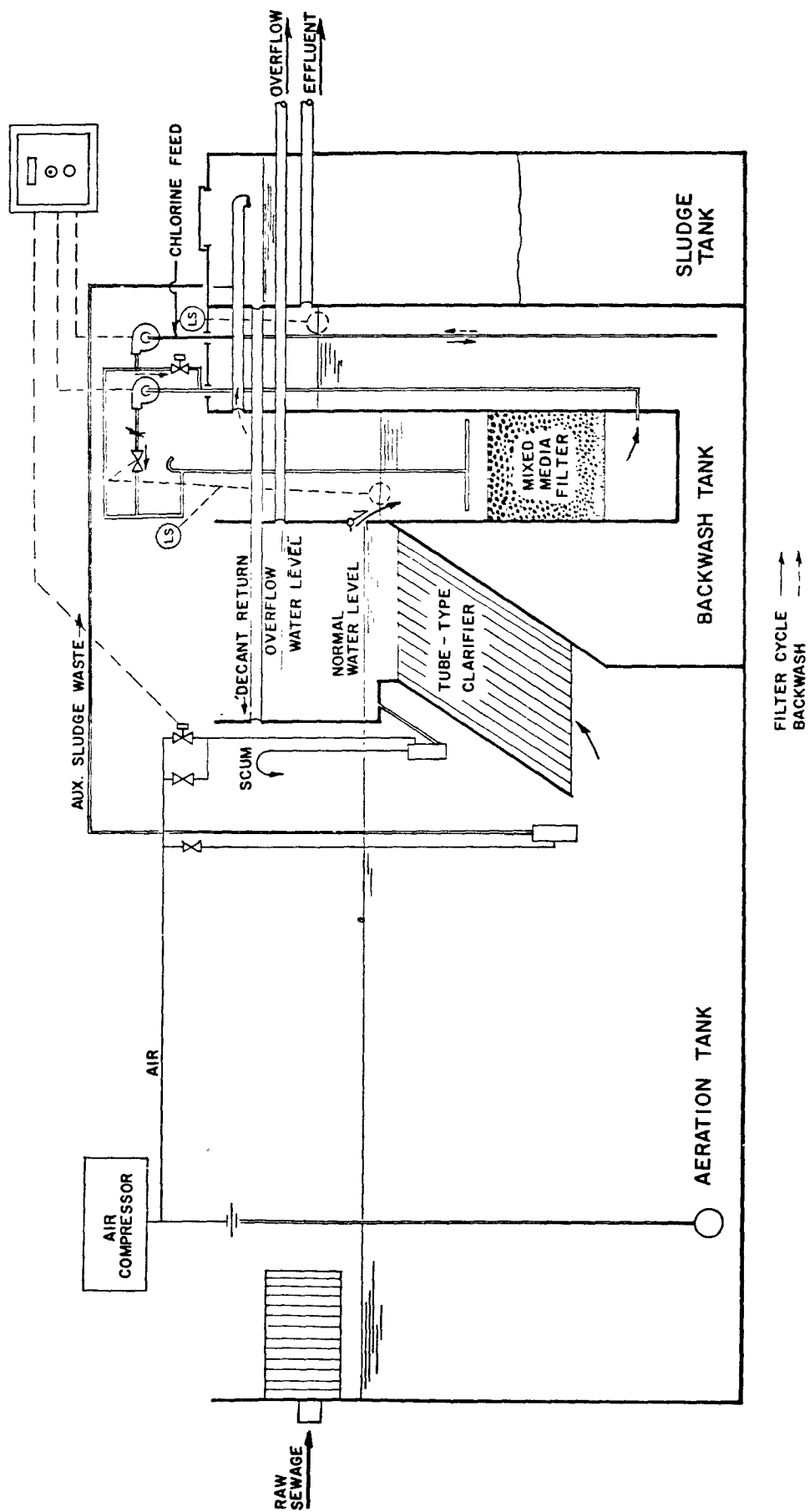
FIGURE 12

FLOW SCHEMATIC OF 3000 GPD PILOT
PLANT USED IN EVALUATION OF STEEPLY
INCLINED TUBES FOR CONTINUOUS
SEPARATION OF MIXED LIQUOR

by 10 feet. A 20,000 gpd plant providing 12 hours aeration, tube settling, mixed-media filtration, backwash storage and chlorine contact, and sludge storage within one package module would occupy only 25 feet by 10 feet. Both plants have a depth of 10 feet. The dimensions of the extended aeration plant do not include the area required by sludge storage facilities provided, nor the area required for chlorine contact. Thus, the integration of high-rate sedimentation and filtration technology enable the design of very compact treatment systems which provide a higher degree of reliability and a higher degree of effluent quality than conventional systems while occupying less space.

Another system into which the steeply inclined, self-cleaning tube settler has been integrated is shown in Figure 13. The quality of effluent obtainable from a trickling filter plant is often limited by the failure of the filter to efficiently remove soluble BOD. Removal of this soluble BOD can be achieved by activated sludge treatment of the trickling filter effluent. The system shown in Figure 13 is designed to accomplish this removal within the confines of the existing secondary clarifier. This is achieved by converting the existing secondary clarifier to an aeration basin and including a steeply inclined tube settler within the existing basin. The very small volume required for solids separation in the steeply inclined tube settler allows both the necessary aeration time and sedimentation time to be provided within the existing basin. A 150,000 gpd plant has been converted to this process and has been in operation for several months. The data from this operation have been reported in detail (4). This operation has shown the complete removal of soluble BOD to be achieved in the modified clarifier which provides 1-1/2 hours of aeration and 10 minutes of settling in the steeply inclined tubes. The suspended solids escaping the inclined tubes in this application have averaged 30-40 mg/l. Subsequent filtration, as provided in the pilot filter shown in Figure 13, reduces the suspended solids and BOD to less than 5 mg/l. The integration of the tube settler into this system allows the efficiency of a trickling filter plant to be upgraded from the norm of 85 percent to more than 95 percent with the only additional area required being that for the relatively small space occupied by a mixed-media filter. Solids in the aeration zone of the modified clarifier have been 2,000-4,000 mg/l during these tests.
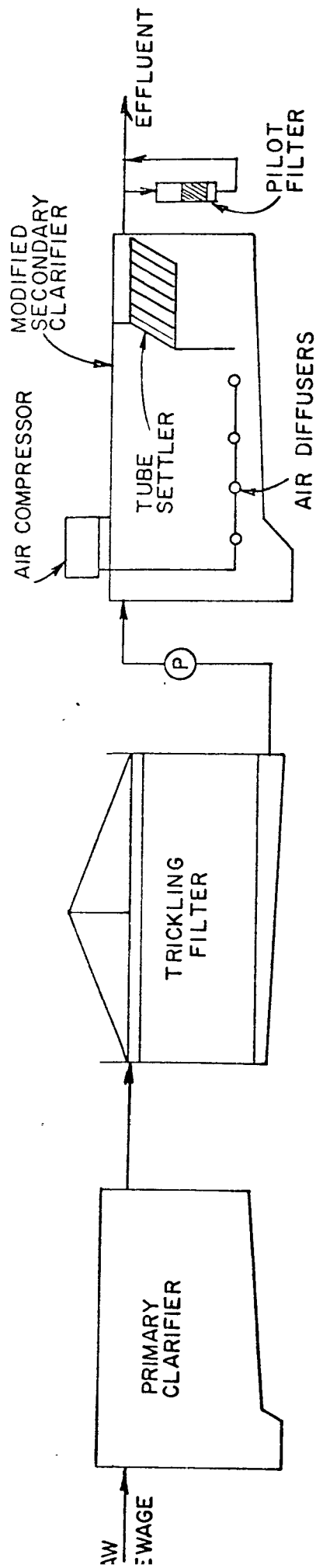
FIGURE 13

SCHEMATIC DIAGRAM OF 150,000 G.P.D.
TRICKLING FILTER PLANT
MODIFIED FOR TESTING OF
UPGRADING TECHNIQUE

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THE DOW CHEMICAL COMPANY

THE DOW PROCESS
FOR
PHOSPHORUS REMOVAL

by

RONALD F. WUKASCH
for the

FWPCA PHOSPHORUS REMOVAL SYMPOSIUM

CHICAGO, ILLINOIS
JUNE 1968

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THE DOW PROCESS FOR PHOSPHORUS REMOVAL

by: Ronald F. Wukasch
    Environmental Engineer
    The Dow Chemical Company
    Midland, Michigan

Introduction

Late in 1966, The Dow Chemical Company instituted a program which sought to develop a phosphate removal process compatible with both the activated sludge and trickling filter processes. For several reasons, a chemical rather than a biological approach to phosphoru removal was selected for study, not the least of which is the reservoir of strength in chemical processes within the company.

Process Description

Chemical phosphorus removal techniques involve the conversion of soluble phosphorus into an insoluble form by precipitation or adsorption. This can be accomplished by the addition of metallic salts to the wastewater to form slightly soluble metal phosphate compounds, or by increasing the pH of the wastewater with lime or other strongly basic materials to cause the formation of insoluble calcium phosphate salts.

Precipitation of soluble phosphorus to the insoluble form, however, does not complete a phosphorus removal process. The resulting precipitates are extremely fine and well suspended. Without further treatment, these fine phosphorus precipitates resist removal by sedimentation. Accordingly, an effective chemical phosphorus removal process must have two essential steps.

1. The conversion of soluble phosphates into an insoluble, suspended form, and
2. The permanent removal of the suspended phosphate from the waste stream.

The first step, the conversion of soluble phosphates to insoluble, suspended phosphates, is accomplished with the addition of small quantities of metallic salts. (Figure 1) Ferrous, ferric, and aluminum salts are all effective.

The second step is the removal of the insoluble phosphorus from suspension. Under normal conditions, only 10 to 15 percent of these extremely fine solids can be removed by sedimentation. However, these suspended metallic phosphates can be flocculated with suitable organic polyelectrolytes. For example, the suspended phosphate particles formed by the addition of ferric chloride or sodium aluminate, are readily coalesced into large, well-settling flocs with anionic organic polyelectrolytes. The completeness of flocculation or solids capture of these systems is excellent and the settling rate of the flocculated material is rapid.

The colloidal suspension formed by the addition of a small quantity of ferrous chloride requires modification by the addition of a strong base before it can be readily flocculated with organic polyelectrolytes. The transient strong base alkalinity modifies the colloidal properties of the system. It is not a pH associated phenomena.

Typically the chemical requirements for the ferrous chloride system are 10 to 25 mg/l $FeCl_2$ as Fe, 30 to 40 mg/l of strong base alkalinity as $CaCO_3$ (24 to 32 mg/l of NaOH or 17 to 22 mg/l CaO), and 0.3 to 0.5 mg/l of Purifloc® A-23, a high molecular weight anionic polyelectrolyte.

Application

The three choices for plant-scale use of this chemical system
are in primary sedimentation, secondary clarification, or
a tertiary sedimentation system. Use of the process in primary
sedimentation promises the greatest benefits for the reasons
described in Table 1.

Plant Scale Data

Grayling, Michigan - The process was demonstrated full-scale
at two Michigan sewage treatment plants during the summer of 1967.
The first of these studies was conducted over a period of three
months at the Grayling, Michigan, waste treatment plant which
provides primary treatment for a flow of about 0.3 MGD. As shown
in Figure 2, the average chemical dosages were 15 to 25 mg/l
$FeCl_2$ as Fe, 30 to 50 mg/l NaOH as $CaCO_3$, and 0.3 mg/l of
anionic polyelectrolyte. During the trial the plant was heavily
overloaded hydraulically due to the seasonal influx of tourists
and a National Guard summer encampment. In addition, no attempt
was made to proportion the chemical feed rates to the variations
in flow at the plant, and consequently chemical dosages were
either above or below optimum most of the time. Despite these
adverse circumstances, total phosphorus removal ranged between
60 and 80 percent with a mean of 72 percent as shown in Table 2.
Suspended solids removal was increased by 27 percent to an average
of 78 percent. BOD removal increased from 40 percent to 58 percent
during this study. The process is being installed more permanently
at this plant and will be operational as of June 1968.

Table 1

BENEFITS OF PHOSPHATE REMOVAL IN PRIMARY TREATMENT

*Process applicable in both activated sludge and trickling
 filter plants.

*Secondary biological treatment provides "polishing" for further
 phosphate removal.

*Increased removal of SS and BOD in primaries.

*Reduced loading to secondary

    *Improved final effluent quality

    *Reduction of waste activated sludge or
     trickling filter humus

*Increased ratio of primary sludge to secondary
 sludge

    *Higher concentration digester feed

    *Improved sludge dewatering

    *Lower total volumes of sludge to be handled

Table 2

Grayling, Michigan Chemical Treatment Studies

Summer 1967

	Influent, mg/l		Effluent, mg/l		% Removal		N
	Avg.	S	Avg.	S	Avg.	S	
Suspended Solids							
No chemicals	187	116	83	23	49.7	18.9	38
With chemicals	223	67	50	13	76.8	7.3	28
BOD_5							
No chemicals	181	57	98	21	43.8	12.9	38
With chemicals	179	26	74	10	57.8	9.6	28
Total Phosphate							
mg/l as P:							
No chemicals					10-20		0
With chemicals	15.8	5.1	4.3	1.3	72.3	8.0	28

S = Standard deviation

n = Number of samples (composited 8 A.M. to 4 P.M., three days per week)

Lake Odessa, Michigan - The second plant scale test was performed over a three-month period at Lake Odessa, Michigan, with a flow of 0.5 MGD (Figure 3). Here the treatment facilities included primary sedimentation, a trickling filter, and sludge digestion with the digester supernatant returned to the primary clarifiers. Chemical dosages averaged the same as used in the Grayling study with no attempt made to proportion chemical feed rate to plant flow. With chemical treatment, total phosphorus removal ranged from 75 to 93 percent with a mean value of 82 percent. (Table 3) Overall plant suspended solids removal and BOD removal were also substantially improved.

Mixing and flocculation were much less than required for maximum process efficiency, and physical modifications of the plant could be expected to substantially increase the effectiveness of the process. Permanent installation of the process which requires modifications is currently under design.

The phosphorus concentration in the returned digester supernatant decreased during the period of this study. Initial supernatants contained 100 to 200 mg/1 of phosphorus. There was an average of 23 mg/1 total phosphorus in the supernatant during the latter portions of the chemical treatment. It was apparent from this that phosphorus release was not occurring during digestion Additional proof was obtained when a crystalline phosphate precipitate, vivianite $(Fe_3PO_4)_2 \cdot 8H_2O$ was identified in the digested sludge from Lake Odessa and also in the Grayling undigested primary sludge by x-ray diffraction techniques. (Table 4) The quantitative elemental analysis by x-ray fluorescence of these two sludges is displayed in Table 5.

Table 3
Lake Odessa, Michigan Chemical Treatment Studies
Summer 1967

	Influent Avg.	S	Primary Effluent Avg.	S	% Primary Removal Avg.	S	Final Effluent Avg.	S	% Overall Removal Avg.	S	N
Suspended Solids, mg/l											
No chemicals	370	143	140	50	62	13	77	27	79	8	22
With chemicals	478	169	99	26	79	9	55	31	89	6	22
BOD_5, mg/l											
No chemicals	360	241	249	181	31	21	141	105	61	15	22
With chemicals	191	60	117	50	39	19	39	27	80	16	22
Total Phosphate											
mg/l as P											
With chemicals	10.8	5.2	3.1	1.0	71.0	16.6	1.7	0.4	82.5	14.4	22

S = Standard deviation

n = Number of 24 hour composite samples

Table 4

SLUDGE ANALYSES : PHOSPHATE REMOVAL

Ferrous Chloride - Hydroxide - Polyelectrolyte Process

Qualitative Analysis: (X-ray Diffraction)

Grayling Primary Sludge:
Major Const: Vivianite; $Fe_3(PO_4)_2 \cdot 8H_2O$

Minor Const: Quartz; SiO_2

Unidentified line at 16 Å

Lake Odessa Mixed Primary and Humus:
Major Const: Vivianite; $Fe_3(PO_4)_2 \cdot 8H_2O$

Minor Const: Quartz; SiO_2

Table 5

SLUDGE ANALYSES : PHOSPHATE REMOVAL

Ferrous Chloride - Hydroxide - Polyelectrolyte Process

Quantitative Analysis: (X-ray Diffraction)

Element	Percent	
	Grayling Primary	Lake Odessa Mixed
Pb	-	-
Ba	0.12	.02
Sn	.01	-
Cd	-	-
Ag	.01	-
Sr	-	-
Zn	.23	.24
Ca	.14	.1
Fe	20.2	23.6
Mn	.08	.06
Ti	.11	.1
Ca	8.1	6.0
K	.4	.8
Cl	1.1	2.1
S	3.1	3.0
P	5.1	6.8
Si	.4	4.0
Al	.8	.8

Since the use of this process in primary treatment significantly increases suspended solids removal from the raw sewage, the quantity of primary sludge produced in a plant is increased by additional sewage solids as well as phosphate solids. Thus, it is difficult to tell directly how much additional non-volatile sludge solids can be directly attributable to the chemical process. However, by estimating that 5 mg/l of additional phosphorus will be removed as vivianite with the remaining iron being removed as the hydroxide, it can be estimated that approximately 300 pounds of additional inorganic sludge solids will be produced per million gallons. The sodium or calcium added to the waste as alkalinity remains in solution.

The plant scale studies at Grayling and Lake Odessa were conducted in cooperation with the Michigan Department of Public Health Copies of the joint reports that have issued from these studies are available. [1,2]

Chemical and Physical Requirements
Metal Requirements - The removal of total phosphorus from the waste stream is limited by the thoroughness to which the soluble phosphorus has been converted to insoluble phosphorus (Step 1) The degree to which soluble phosphorus is insolubilized obviously depends on the metal dosage The quantity of the metals which must be added to achieve certain results in terms of final ortho-phosphate concentration in the waste stream can be predicted as shown in Figures 4 through 7 These data represent laboratory studies at 11 different waste treatment plants in Michigan and Ohio with varying degrees of industrial components. These curves show that the amount of metal that must be added per unit of phosphorus removed from solution regresses is a function of the desired final ortho-phosphate concentration. This functional relationship does not describe the removal mechanism, since both adsorption and precipitation phenomena fit the same model.

However, these relationships are useful in predicting the metal
dosages required to achieve certain results in phosphorus removal.

Flocculation Requirements - Having accomplished the insolubili-
zation of the soluble phosphorus present in the waste stream,
the problem now becomes removing it from suspension (Step 2).
As stated previously, these extremely fine colloidal precipitates
settle poorly and require chemical flocculation to permit removal
by sedimentation. This process requirement of flocculation
is shown quantitatively in Figures 8 through 11. The experimental
procedure used to generate these data is a modified jar test con-
ducted on one liter samples in 1500 ml beakers on a Phipps and
Byrd gang stirrer. A 5 minute final settling period, under dynamic
(stirred) conditions, was used to represent sedimentation tank
performance.

In each of these systems, chemical flocculation of the precipitated
phosphorus is required to achieve satisfactory total phosphorus
removal. The alum and polyelectrolyte system has shown a highly
erratic response on certain wastes, even with alkalinity adjust-
ments.

Reaction Time Requirements - There is a reaction time criticality
between the addition of the metallic ions and polyelectrolyte
addition.

As shown in Figures 12 and 13 for the case of ferrous iron and
sodium aluminate, a minimum intevening mixing period of approx-
imately 4 minutes is required between inorganic chemical addition
and organic polyelectrolyte addition. This type of response
is also the case for ferric iron and alum. Apparently, this
reaction time is not necessary to the insolubilization of the
dissolved phosphorus since the ortho-phosphate disappearance is
almost immediate upon the addition of the metal. Presumably a

change in the colloidal system is occurring during this intervening mixing period which results in a suspension much more responsive to anionic polyelectrolyte flocculation.

Table 6 presents a summary of the operational requirements, based upon the preceding studies, that must be met to successfully apply this process for phosphate removal.

Cost of Process
A cost summary of the various modifications of the process is shown in Table 7. The basis for these cost figures is shown in the appendix.

The chemical costs quoted are representative, fair price per pound figures. Local conditions, shipping distances, and other factors will affect these nominal prices. In addition, variations in the quantity of phosphorus and response of the waste stream from city to city will alter chemical requirements.

The total cost of the capital equipment is not very sensitive to the size of the plant. The capital costs shown are based on the equipment required for a plant treating between 50 and 100 MGD. One key piece of equipment is an automatic ortho-phosphate analyzer used to control the metal salt dosage on a continuous basis. The iron or aluminum dose must be at correct ratio to produce the desired residual ortho-phosphate. Too low a dosage reduces the effectiveness of phosphate removal; too high a dosage is uneconomical and it will result in iron or aluminum carry-over in the plant effluent.

A modest reduction in overall sludge handling costs can be expected. Although some additional inorganic solids are being produced in the process, the reduction of secondary sludge due to improve primary sedimentation efficiency represents a real cost savings in terms of digester volume and dewatering costs.

Table 6

SUMMARY OF PHYSICAL REQUIREMENTS
PHOSPHATE REMOVAL PROCESS

I.   Add metal salts to the raw sewage with thorough mixing.

    A.   With $FeCl_2$, add base not less than 10 seconds later.

II.   Allow reaction to proceed for a minimum of five minutes.

III.   Add A-23 organic polyelectrolyte.

    A.   Flash mix 20 to 60 seconds.

IV.   Mechanical or air flocculation; 1 to 5 minutes

V.   Gentle delivery of flocculated sewage to sedimentation tanks.

Table 7

PHOSPHATE REMOVAL PROCESS COSTS

(In Dollars per Million Gallons)

Chemical System	Chemical Costs				Capital Costs	Operating	Total Cost per MG
	Metal	Alkalinity	A-23	Total			
$FeCl_2$:CaO:A-23*	12.30	1.67	5.65	19.62	0.59	2.46	22.67
$FeCl_2$:NaOH:A-23	12.30	7.00	5.65	24.95	0.51	1.26	26.72
$FeCl_2$:A-23	38.80		5.65	44.45	0.38	1.26	46.09
$Na_2Al_2O_4$:A-23	48.00		5.65	53.65	0.39	1.26	55.30

*Purifloc A-23 by The Dow Chemical Company

CONCLUSIONS

1. A two step process consisting of
   1) Converting soluble phosphate to suspended phosphate
      with iron or aluminum salts.
   2) Removal of the suspended phosophate and other solids
      by chemical flocculation and sedimentation

   has been demonstrated to be an effective method for
   phosphorus removal.

2. Practice of the process in primary sedimentation results
   in greatly improved suspended solids and BOD removal in
   primary treatment with corresponding plant-wide benefits.

3. With minor physical modifications, the process can be
   integrated into virtually any waste treatment plant.

4. The total cost of the process for various chemical
   systems can range from $25 to $60 per million gallons
   treated.

5. The system using ferrous chloride, alkalinity and an
   anionic polyelectrolyte is the most economical variation
   of the process with a total cost of less than $30 per
   million gallons treated.

REFERENCES

1. "Studies on Removal of Phosphates and Related Removal of
    Suspended Matter and Biochemical Oxygen Demand at Grayling,
    Michigan, March-September 1967," Otto Green, Fred Eyer,
    and Donald Pierce, Division of Engineering, Michigan
    Department of Health and The Dow Chemical Company.

2. "Studies on Removal of Phosphates and Related Removal of
    Suspended Matter and Biochemical Oxygen Demand at
    Lake Odessa, Michigan, May-October, 1967." Wastewater
    Section, Division of Engineering, Michigan Department of
    Health and The Dow Chemical Company.

APPENDIX

Chemical Costs

I.   Ferrous Chloride: $FeCl_2$

Assume 4 mg/l orthophosphate as P reduced to 0.3 mg/l.
Requires 4.0 mg Fe(II)/mg P.

4.0 mg Fe/mg P x 3.7 mg/l P = 15 mg/l Fe(II)

$15 \times \frac{127}{56}$ = 37 mg/l $FeCl_2$

= 308 lbs/MG $FeCl_2$

@ 4¢/lb = $12.30/MG

II. Ferric Chloride: $FeCl_3$

Assume 4 mg/l orthophosphate as P reduced to 0.3 mg/l.
Requires 10.9 mg Fe(III)/mg P.

10.9 mg Fe/mg P x 3.7 mg/l P = 40 mg/l Fe(III)

$40 \times \frac{162.5}{56}$ = 116 mg/l $FeCl_3$

= 970 lbs/MG

@ 4¢/lb = $38.80/MG

III. Sodium Aluminate:

Assume 4 mg/l orthophosphate as P reduced to 0.3 mg/l.
Requires 6.2 mg Al/mg P.

6.2 mg Al/ml P x 3.7 mg/l P = 23 mg/l Al

$23 \times \frac{100}{24\% \ Al}$ = 96 mg/l $Na_2Al_2O_4 \cdot 3H_2O$

= 800 lbs/MG

@ 6¢/lb = $48.00/MG

IV.   Sodium Hydroxide: NaOH
      28 mg/l NaOH = 234 lbs/MG
              @ 3¢/lb = $7.00/MG

V.   Lime: CaO
      20 mg/l CaO = 167 lbs/MG
              @ 1¢/lb = $1.67/MG

VI.   Dow Purifloc A-23
      0.5 mg/l = 4.17 lbs/MG
      @ $1.35/lb. = $5.65/MG
      (Truck load quantities)

Capital Equipment Costs
(Treating 50 to 100 MGD)

I. INORGANIC CHEMICAL STORAGE AND HANDLING

		$FeCl_2$ or $FeCl_3$	NaOH	CaO or $Na_2Al_2O_4$
A)	GENERAL (Total)	$ 4,975	$ 3,595	$ 4,045
B)	PIPING (Total)	6,620	4,045	660
C)	EQUIPMENT (Total)	16,000	8,700	25,000
D)	ELECTRICAL REQUIREMENTS (Total)	900	2,100	400
	TOTAL DIRECT COST	$28,495	$18,440	$30,105
	TOTAL including Overhead, Profit, and Engineering (30%)	$37,200	$24,000	$39,200
	Cost per million gallons amortized 10 years @ 5% for 50 MGD	$0.20	$0.13	$0.21

*Includes electric motors

II. AUTOMATIC CONTROL EQUIPMENT

A) Automatic orthophosphate analyzer and recorder. Automatic addition of metal proportional to orthophosphate concentration and plant flow. Automatic addition of alkalinity and polyelectrolyte proportional to plant flow. Chemical inventory controls and alarms.

Estimated cost = $32,000

Cost per million gallons
amortized 10 years @   = $0.18/MG
5% for 50 MGD

Operating and Maintenance Costs
(50 MGD Plant)

I. MANPOWER

    A. Technical:

    Assume 1 man full time

$$\frac{\$12,000}{50\ MGD \times 365\ Days} = \qquad \$0.66/MG$$

    B. Operator and Maintenance

        1. $FeCl_2$:NaOH:A-23 system

            $FeCl_3$:A-23 system

            $Na_2Al_2O_4$:A-23 system

        Assume 1 man shift per day

$$\frac{\$30}{50\ MGD} = \qquad \$0.60/MG$$

        2. $FeCl_2$:CaO:A-23 system

        Assume 3 man shifts per day

$$\frac{\$90}{50\ MGD} = \qquad \$1.80/MG$$

FIG. 1

PHOSPHATE REMOVAL PROCESS

RAW WASTE : CONTAINS SOLUBLE & SUSPENDED PO_4

STEP 1. CONVERT SOLUBLE PO_4 TO SUSPENDED PO_4
(Add metal ion – Fe, Al, etc.)

STEP 2. REMOVE SUSPENDED PO_4 & OTHER SOLIDS
(Flocculation with polyelectrolytes followed by sedimentation)

PO_4 & OTHER SUSPENDED SOLIDS REMOVED
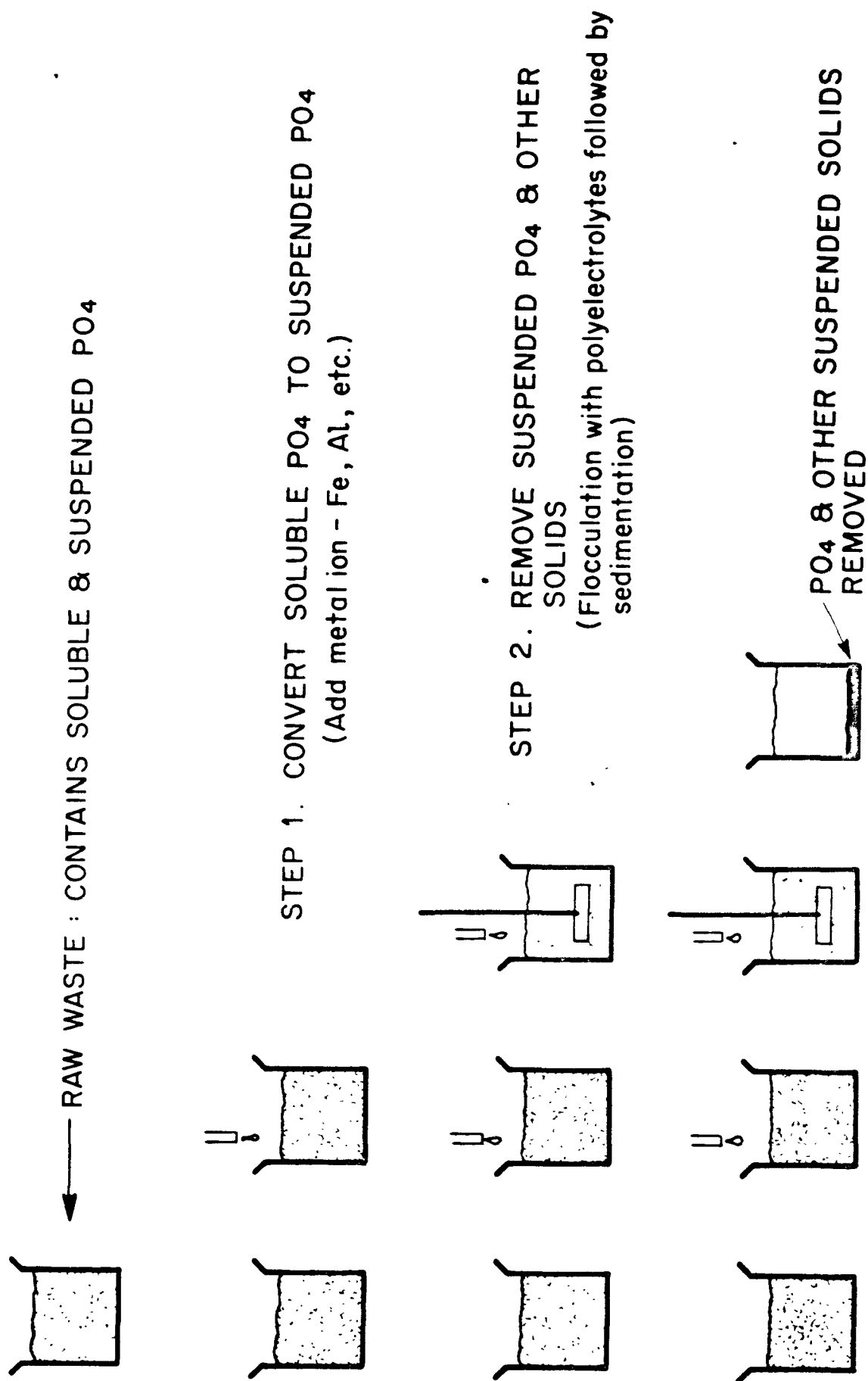
FIG 2
PHOSPHATE REMOVAL PROCESS
GRAYLING, MICHIGAN

PURIFLOC A-23 0 3mg/l

FERROUS CHLORIDE
18mg/l Fe

ALKALINITY
30mg/l NaOH

FLOCCULATOR (?)

PLANT INFLUENT

PUMP
STATION

SETTLING
TANKS

Cl₂

CHLORINE
CONTACT
TANK

Au SABLE RIVER
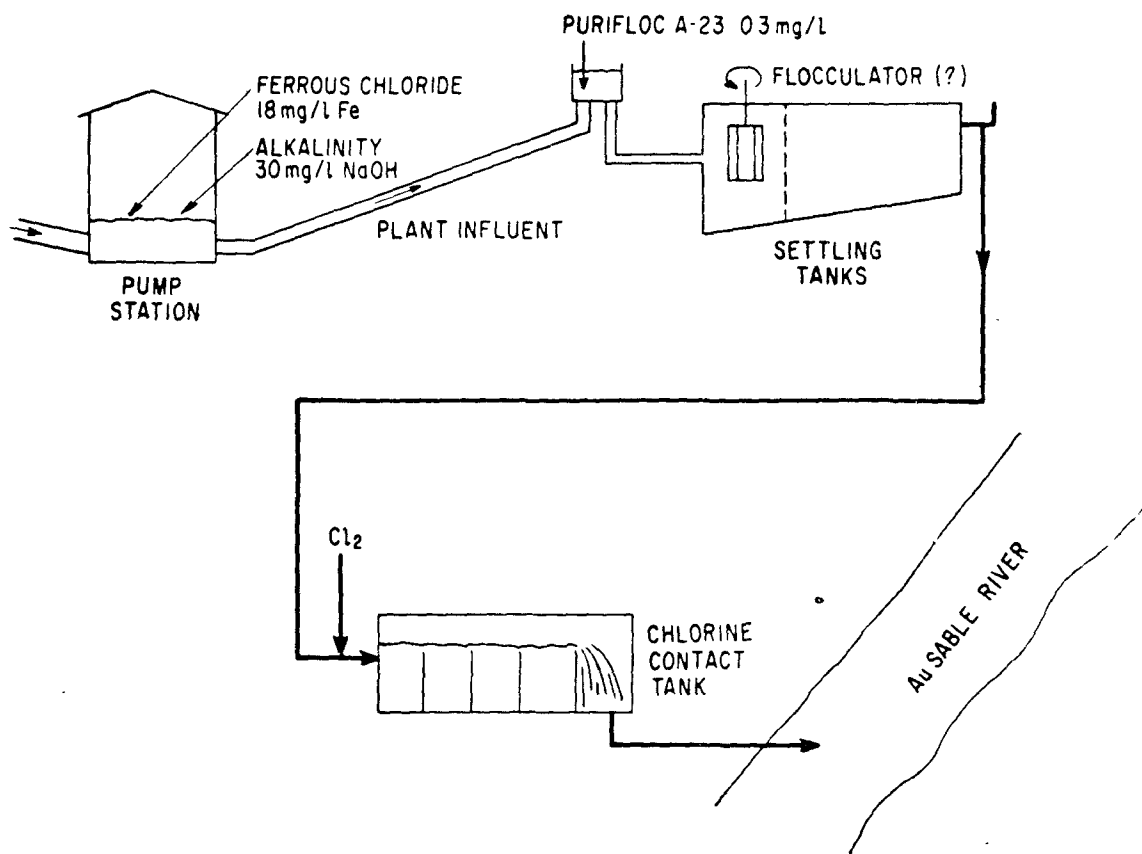
FIG 3
PHOSPHATE REMOVAL PROCESS
LAKE ODESSA, MICHIGAN
1967

FERROUS CHLORIDE
25mg/l Fe
ALKALINITY
30mg/l
NaOH

PURIFLOC
A-23
0 4mg/l

PRIMARY
SEDIMENTATION

TRICKLING
FILTER

FINAL
CLARIFICATION

PLANT
INFLUENT

PARSHALL
FLUME

FINAL
EFFLUENT

SECONDARY
SLUDGE

PRIMARY SLUDGE

SUPERNATANT
RETURN

SLUDGE
DIGESTER

DIGESTED
SLUDGE
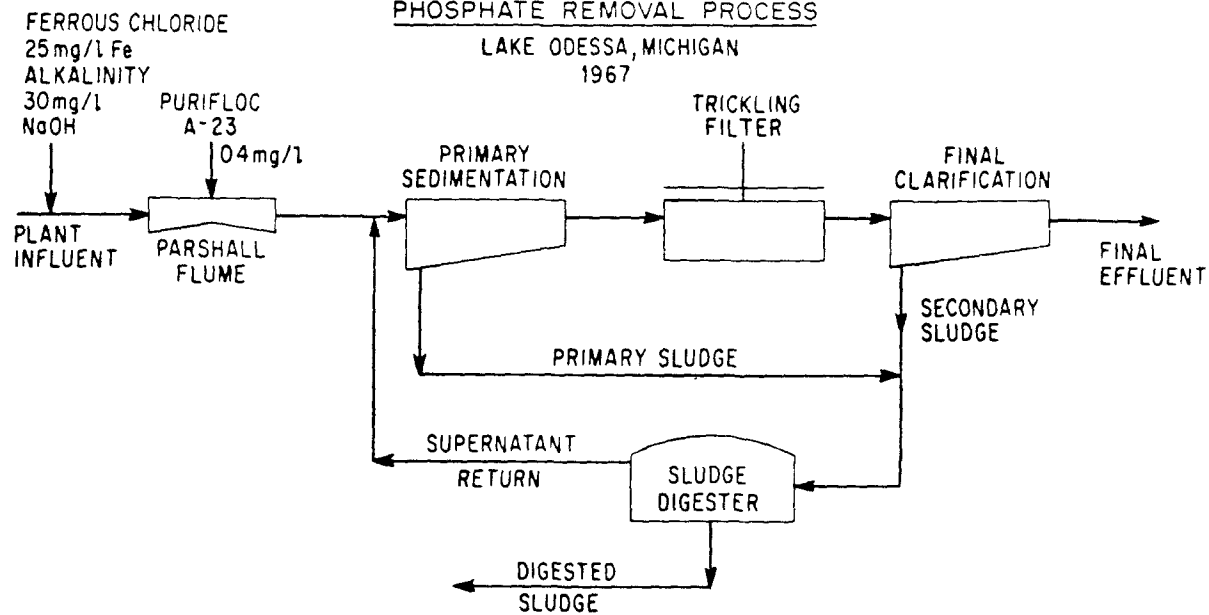
FIGURE 4

INSOLUBILIZATION OF ORTHO PHOSPHATE WITH
FERROUS CHLORIDE FeCl$_2$

FINAL ORTHO PHOSPHATE: mg/l AS PHOSPHOROUS
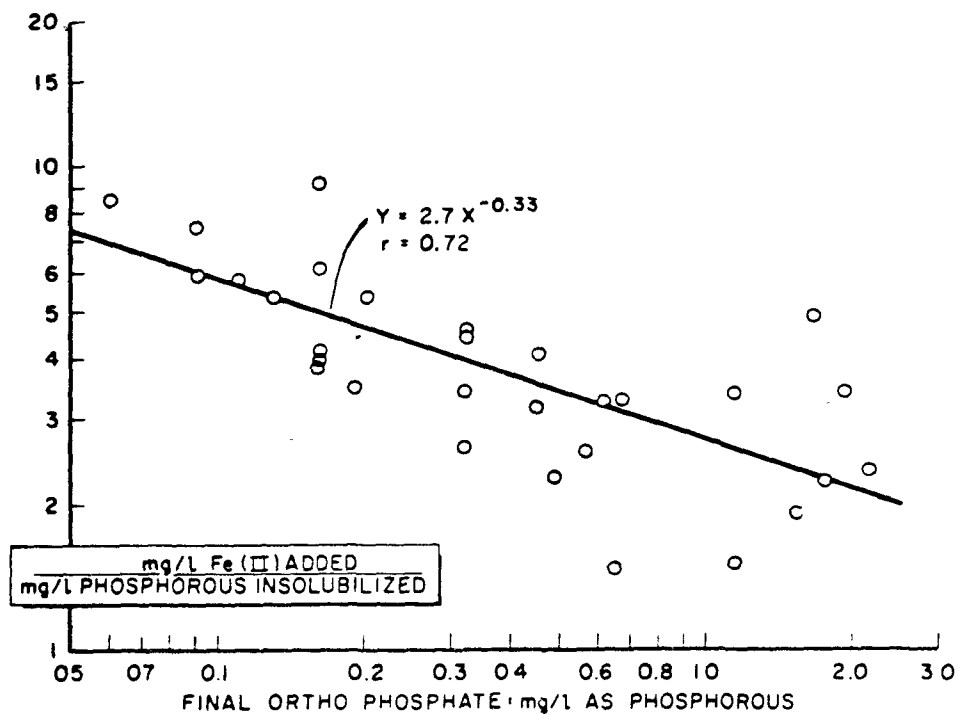
FIGURE 5

INSOLUBILIZATION OF ORTHO PHOSPHATE WITH
FERRIC CHLORIDE FeCl$_3$

FINAL ORTHO PHOSPHATE: mg/l AS PHOSPHOROUS
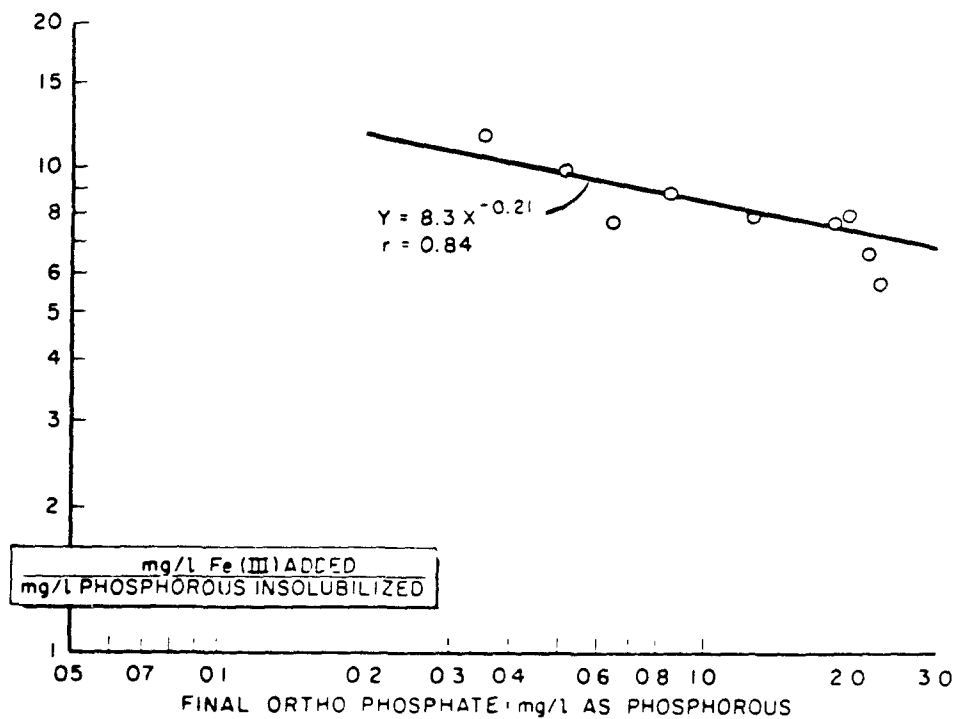
FIGURE 6

INSOLUBILIZATION OF ORTHO PHOSPHATE WITH
ALUM : $Al_2(SO_4)_3 \cdot X \, H_2O$

$Y = 2.3 \, X^{-0.16}$
$r = 0.69$

mg/l ALUMINUM ADDED
mg/l PHOSPHOROUS INSOLUBILIZED

FINAL ORTHO PHOSPHATE : mg/l AS PHOSPHOROUS
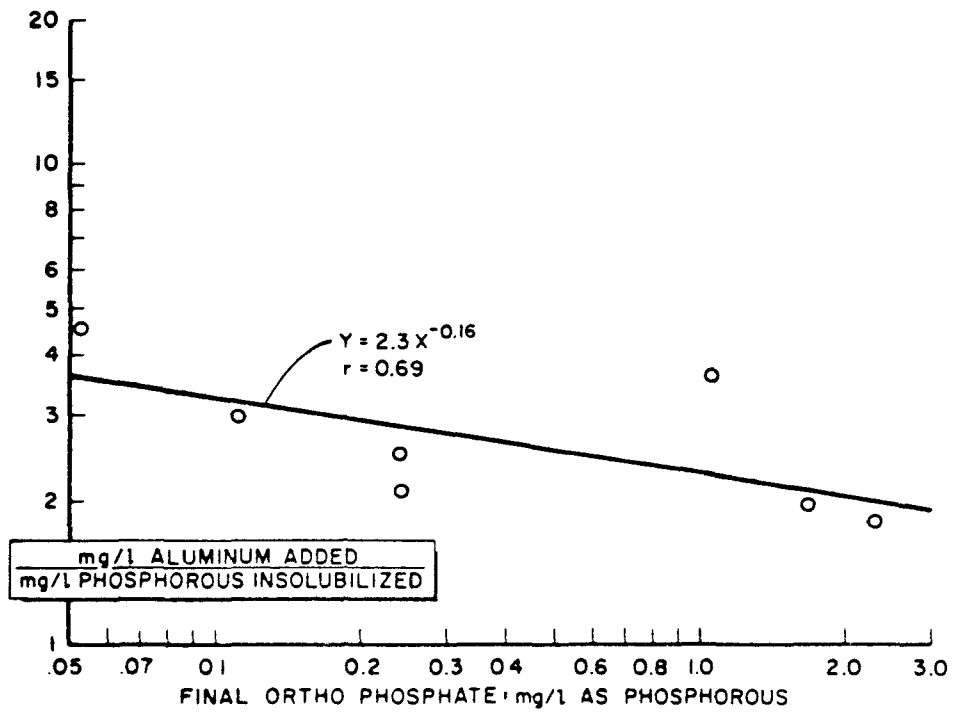
FIGURE 7

INSOLUBILIZATION OF ORTHO PHOSPHATE WITH
SODIUM ALUMINATE : $Na_2Al_2O_4 \cdot 3 \, H_2O$

$Y = 3.7 \, X^{-0.43}$
$r = 0.90$

mg/l ALUMINUM ADDED
mg/l PHOSPHOROUS INSOLUBILIZED

FINAL ORTHO PHOSPHATE : mg/l AS PHOSPHOROUS
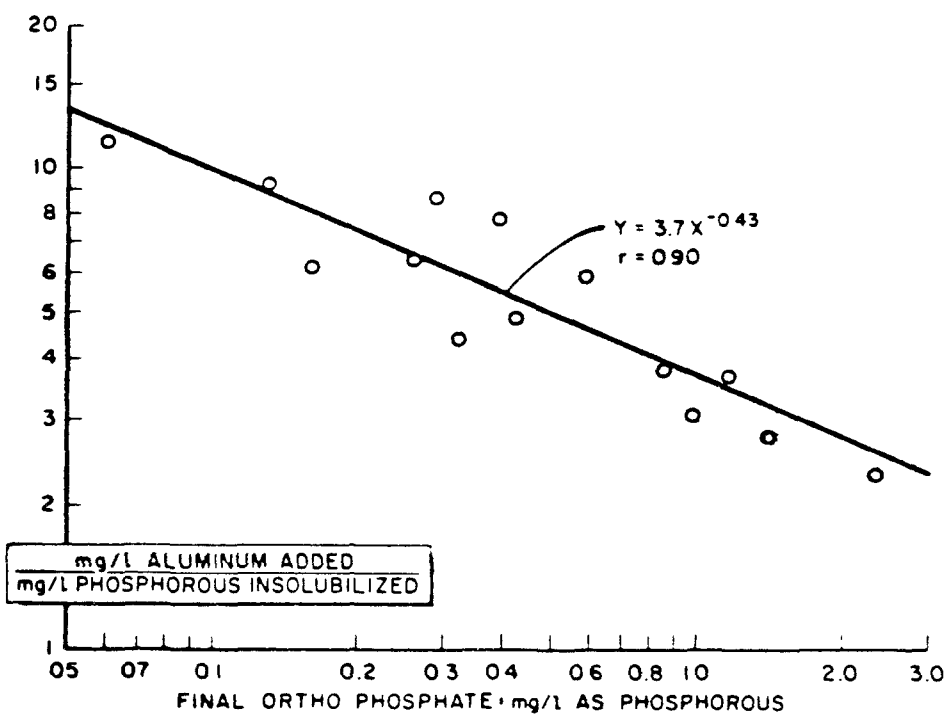
FIGURE 8

PHOSPHOROUS REMOVAL WITH IRON II SYSTEMS

Additional alkalinity plus an organic polyelectrolyte are needed
to remove the suspended phosphates by sedimentation

FERROUS CHLORIDE, FeCl : mg/l Fe(II)
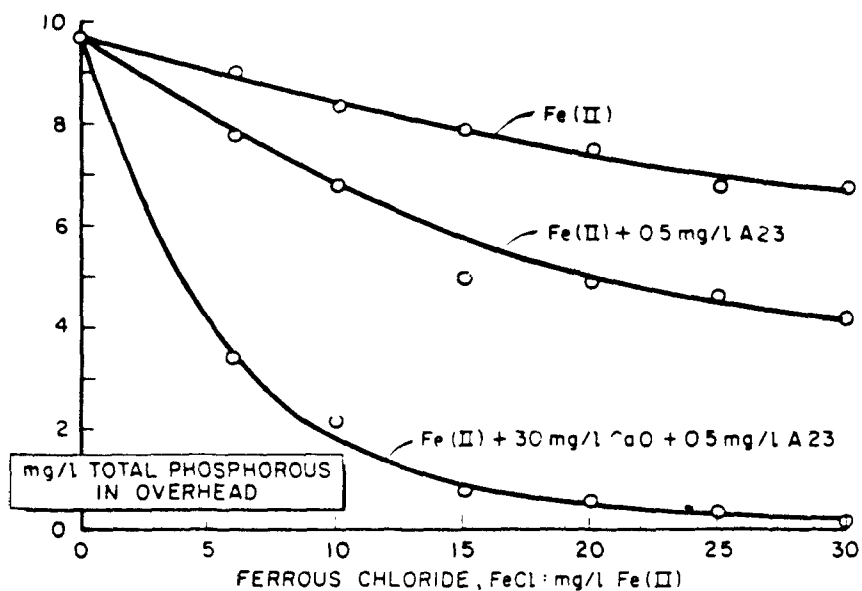
FIGURE 9

PHOSPHOROUS REMOVAL WITH IRON III SYSTEMS

Organic polyelectrolyte is required to flocculate & settle
the suspended phosphates

FERROUS CHLORIDE, FeCl : mg/l Fe(II)
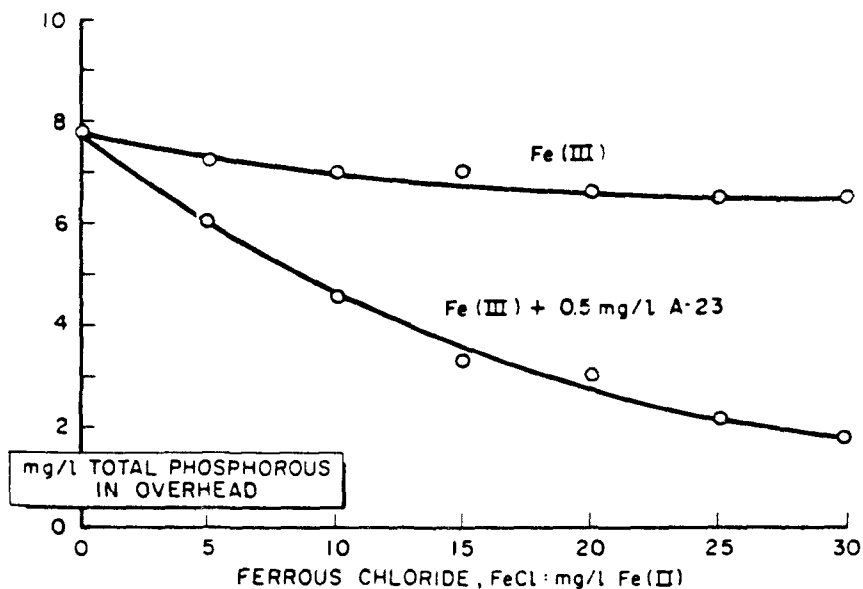
FIGURE 10

PHOSPHOROUS REMOVAL WITH ALUMINATE SYSTEMS

An organic polyelectrolyte is needed to flocculate & settle the insoluble phosphate formed by the addition of sodium aluminate.
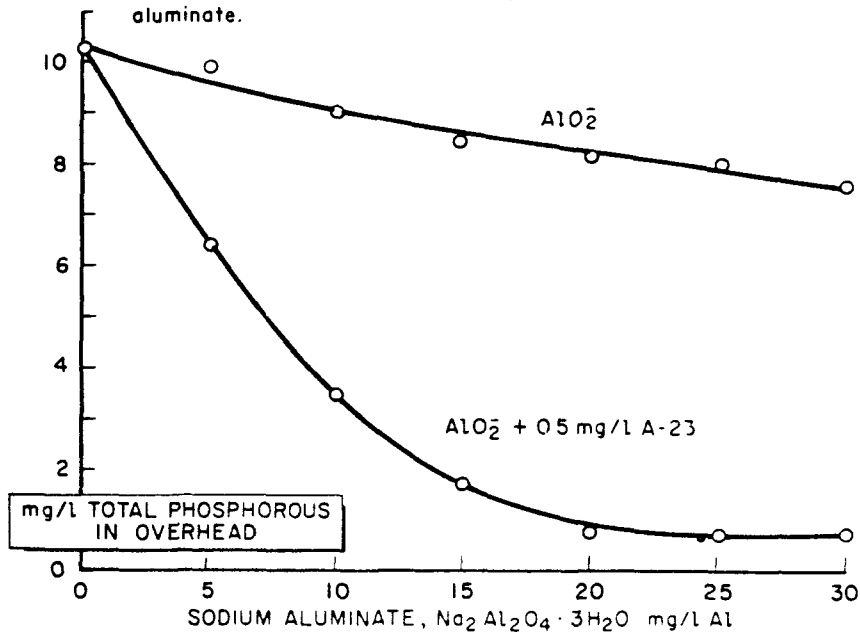
FIGURE 11

PHOSPHOROUS REMOVAL WITH ALUM SYSTEMS

Organic polyelectrolyte is needed to flocculate & settle the suspended phosphates
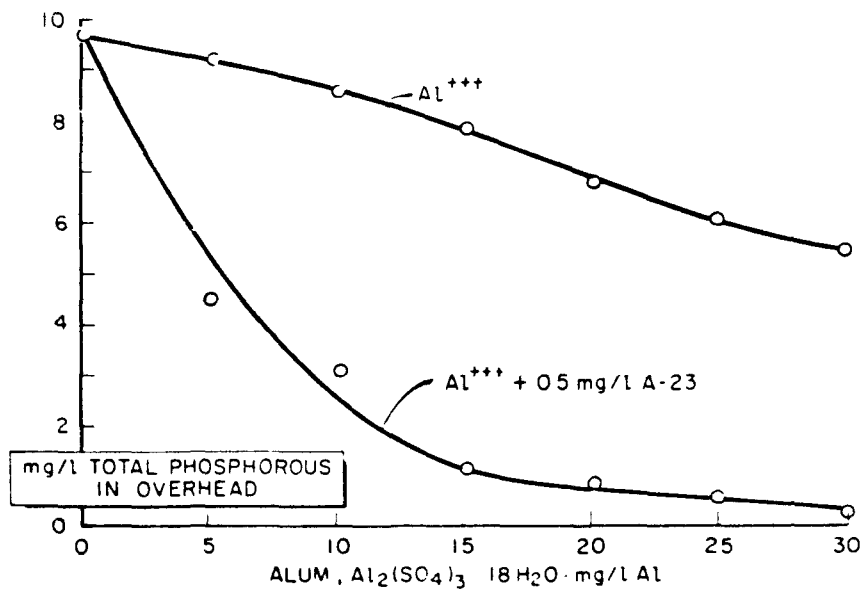
FIGURE 12

REACTION TIME REQUIREMENTS:
IRON (II) SYSTEM

The iron (II) & alkalinity must be added to the waste water at
least 4 minutes before the organic polyelectrolyte

RAW SEWAGE TREATED WITH
FERROUS CHLORIDE:
15 mg/l Fe^{++}

LIME: 30 mg/l CaO

PURIFLOC A-23: 0.5 mg/l

mg/l TOTAL PHOSPHOROUS
IN OVERHEAD

MINUTES BETWEEN INORGANIC CHEMICALS
ADDITION & ORGANIC POLYELECTROLYTE ADDITION
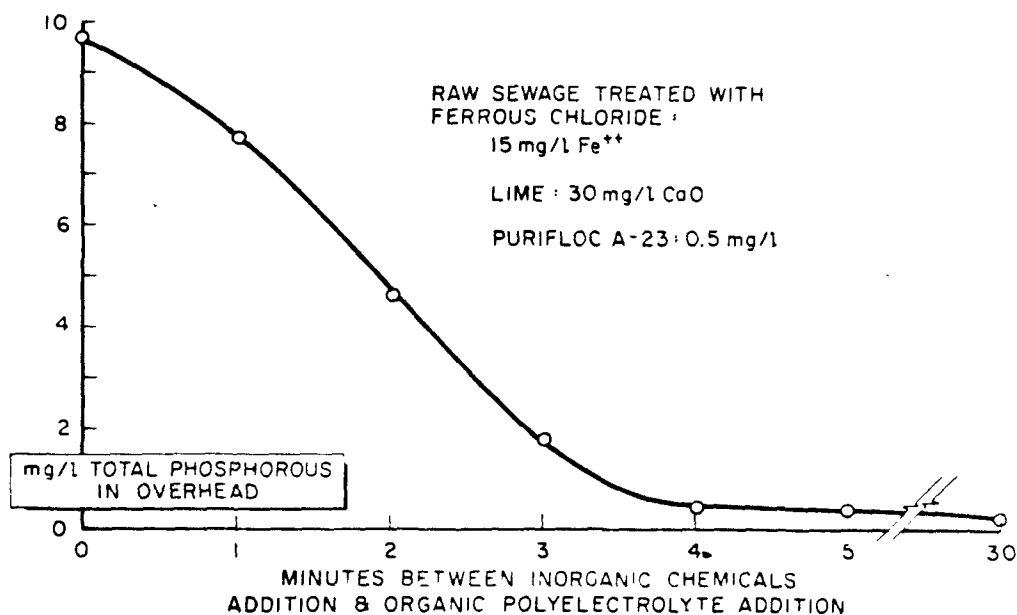
FIGURE 13

REACTION TIME REQUIREMENTS:
ALUMINATE SYSTEM

After the addition of sodium aluminate, a minimum amount of inter-
vening mixing is required before the phosphate containing colloids
can be completely flocculated & settled

RAW SEWAGE TREATED WITH
SODIUM ALUMINATE:
15 mg/l Al
PURIFLOC A-23 0.5 mg/l

mg/l TOTAL PHOSPHOROUS
IN OVERHEAD

MINUTES BETWEEN ALUMINATE
ADDITION & ORGANIC POLYELECTROLYTE ADDITION

FULLER COMPANY / GENERAL AMERICAN TRANSPORTATION CORPORATION

INFILCO PRODUCTS P O BOX 5033. TUCSON. ARIZONA 85703. TELEPHONE 602/623-5401

Mr. Company President, Mayor
or Consulting Engineer
Lake Michigan Area
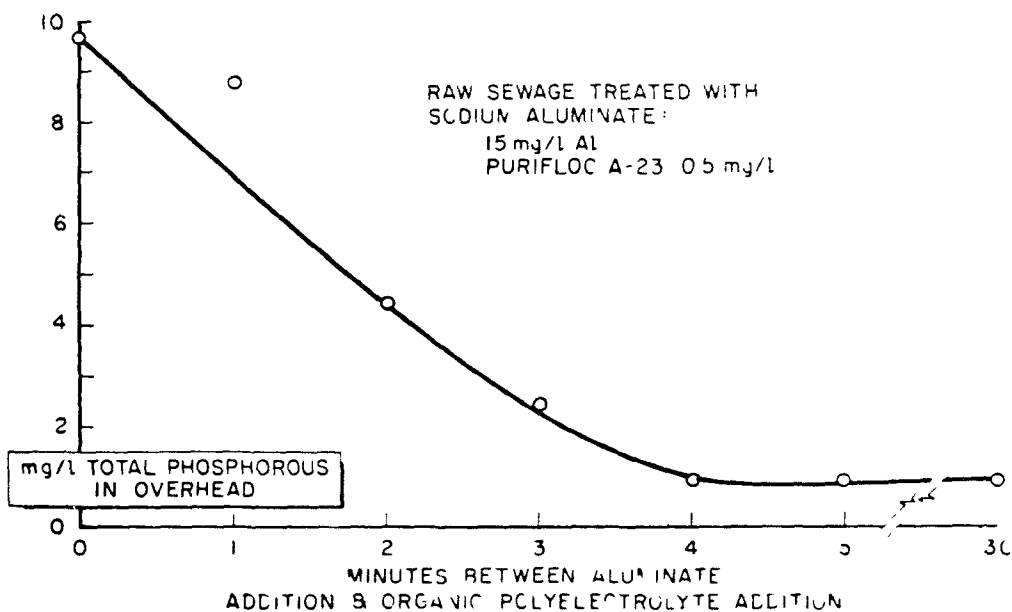
Subject: Water Pollution Problems of Lake Michigan and Tributaries

Dear Sir:

Within the wide scope of corporate activities, many firms (GATX included) and communities now face the necessity of proper housecleaning and adequate wastewater treatment to abate pollution of lakes and streams.

Our Infilco Products group has specialized in water and waste treatment for many years. In putting together this brochure we have attempted to give you an informative "first step" towards wastewater purification. We hope the following information will prompt you to call us in for a discussion — without obligation — of your waste treatment problem, from definition to solution at the least possible cost.

We look forward to hearing from you.

Sincerely yours,

T. E. Meyers
Vice President
Fuller Company

YOUR PROBLEM

"Discharges of untreated and inadequately treated wastes originating in Wisconsin, Illinois, Indiana, and Michigan cause pollution of Lake Michigan which endangers the health and welfare of persons in states other than those in which such discharges originate. This pollution is subject to abatement under the provisions of The Federal Water Pollution Control Act."

ACTION has been recommended by the Lake Michigan Pollution Control Conference, the source of the above statement. For *municipal wastes*, recommended action is to be taken by December 1972 with respect to higher removals of BOD and phosphorus. For *industrial wastes*, the recommended program requires that detailed "action plans" for adequate treatment be developed by September 1, 1968, with treatment facilities to be constructed by December 1972.

The Fuller Company, Infilco Products group, can help you define your specific problem and recommend an optimum

•FULLER / INFILCO SOLUTIONS are based upon:

Cooperation ... with your staff, consulting engineer, and governmental agency in defining your problem and working out an optimum solution.

Analysis ... in the laboratory to produce the necessary information for determining and checking design parameters.

Process Selection ... which will meet regulatory standards yet prove economical in first cost and operating cost.

Scope ... which goes beyond competitive horizons to cover all phases and types of treatment.

Equipment Selection ... of the correct type and design necessary for efficiency and economy of process operation. (Only with a complete line can an unbiased and cost-saving decision be made; and Infilco equipment represents the most complete line in the industry.)

Corporate Experience ... in all phases of water and waste treatment analysis, process know-how and equipment design.

Installation ... facilities are available for erection and start-up on a turn-key basis when such is desirable.

Financing ... which can be explored on a lease basis with Boothe, GATX, an affiliate.

Guarantee ... made in accordance with specification and proposal to assure you of complete satisfaction and compliance.

Follow Up ... continuing service contracts (where desired) to assure optimum operation and minimum maintenance.

SPECIFIC SUGGESTIONS

FWPCA has made the following recommendations for certain municipalities and industries:

Provide adequate secondary biological treatment or its equivalent; advanced waste treatment for phosphate removal; and substantial reduction of nutrients which result in undesirable aquatic growths — by July, 1972.

Substantially eliminate pollution from combined sewers by July, 1977.

For industrial wastes, FWPCA has recommended that:

Detailed action plans for adequate treatment of all industrial wastes be developed within six months (for specified industries).

The Infilco Products Group of the Fuller Company has the capabilities, experience and supply of conventional treatment processes to specific equipment offer to it the development, advanced treatment processes must be used, designed along with detailed continued Infilco has the advanced technique required due to the organization's extensive research and development program in the treatment of water and wastewater.

including phosphate removal, are best accomplished by the "complete mixing" modification of the traditional activated sludge process. The advantages of this approach are incorporated in the modern AERO-ACCELATOR® activated sludge unit.

Additional removals and/or "polishing" of treated effluents are accomplished economically by conventional or in-depth using the GREENLEAF filter control for centralized operation of multiple filters.

Phosphate is also removed by

Efficient and economical removals can be obtained by the *new* Infilco DENSATOR process which embodies the principles of high-density, solids-contact treatment.

THE FOLLOWING SECTIONS ILLUSTRATE AND DESCRIBE THE DETAILS OF THE ABOVE SPECIFIC SUGGESTIONS:

The INFILCO DENSATOR

High-Density, Solids-Contact Treatment Plant
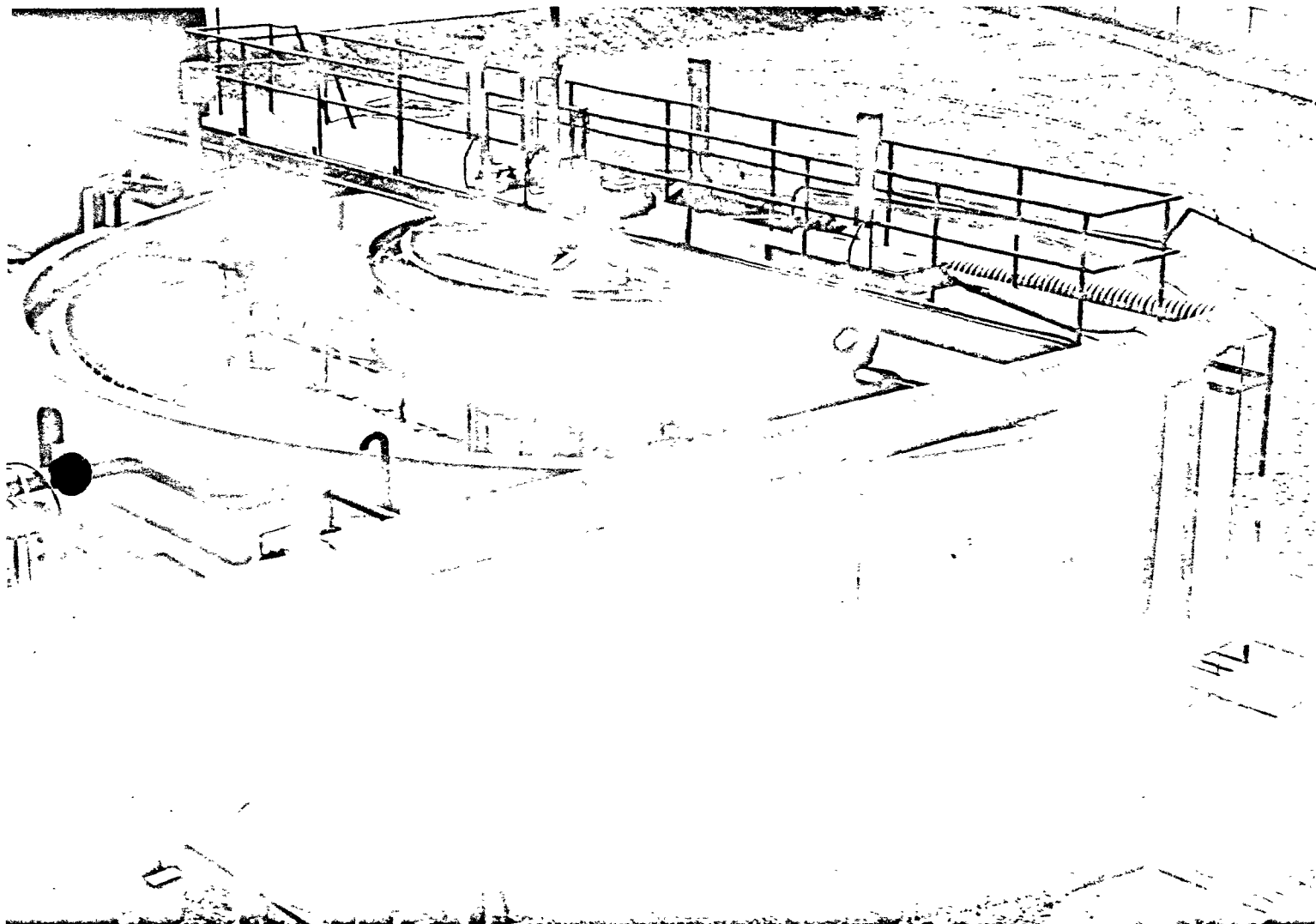
GATX

WHAT'S NEW OR DIFFERENT ABOUT HIGH-DENSITY, SOLIDS-CONTACT UNITS?

Typical High-Rate, Solids-Contact Units

A properly designed high-rate solids-contact treatment unit incorporates within a single unified structure the processes of mixing, coagulation and flocculation, floc conditioning, liquid-solids separation, and automatic sludge removal. Raw water and chemicals are mixed in the presence of previously precipitated solids which have been retained in the treating unit. Chemical reactions are accelerated and colloidal materials easily coagulated since new and smaller particles agglomerate on the surface of old and larger ones. The density of the floc particles and their rate of settling increase in the process. Solids concentrations of 0.5 to 5 grams per liter can be obtained in the mixing zone, with normal concentrations being 1 to 3 grams per liter.

The advantages and economy of the high-rate, solids-contact unit have been proven in both municipal and industrial use, at home and abroad. Since 1934, the Infilco ACCELATOR® high-rate, solids-contact unit has been the most generally accepted design throughout the world.

The DENSATOR Plant

Research and development by Fuller's Infilco product engineers and chemists has established that improved treatment results can be obtained if the solids concentration in the reaction zones is maintained significantly higher than possible with conventional solids-contact units. The DENSATOR plant was designed to provide this process improvement for applications in which it is advantageous.

Normally, the solids concentration in the reaction zones represents the precipitates formed from 5 to 10 reactions; the DENSATOR unit utilizes precipitates formed from up to 50 or more reactions.

The weight of the sludge, after many reactions, increases much more rapidly than its volume. This densification of the sludge is accompanied by rapid settling characteristics — sludge solids of 500 grams per liter have been obtained in the process. Settling rates equivalent to a 4.5 gpm per square foot rise rate have been exceeded with softening types of sludges.

LIME FEED

RAW WATER INLET

PRIMARY REACTION ZONE

TREATED WATER OUTLET

SECONDARY REACTION ZONE

SOLIDS SEPARATION

DRAIN

DENSE SLUDGE

SLUDGE RECIRCULATION
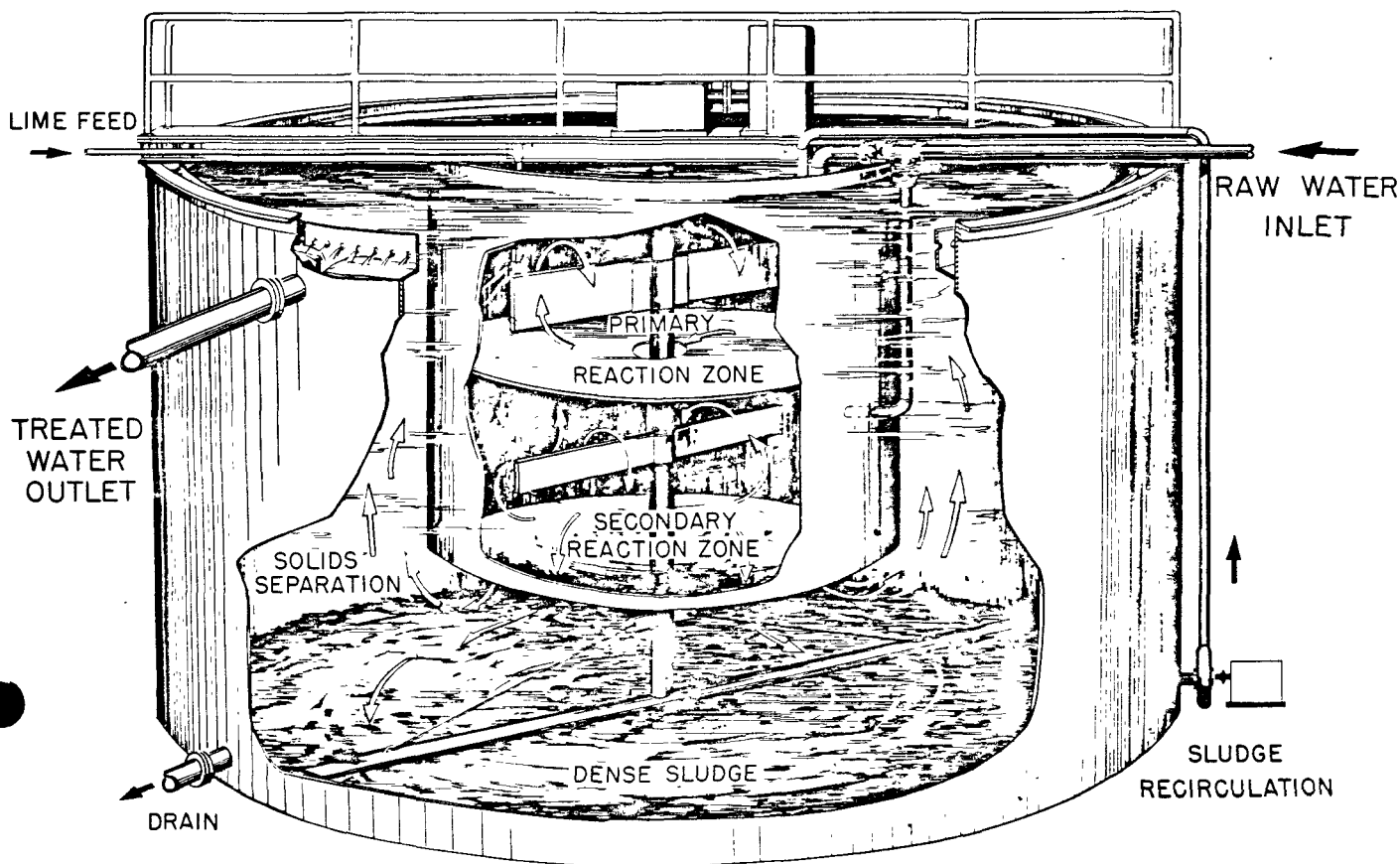
Cutaway drawing of DENSATOR plant
shows treatment zones
and illustrates the process

Advanced Design

High-density, solids-contact treatment requires special modifications of the accepted design of existing solids-contact units. New and improved means of sludge recirculation have been incorporated to replace conventional hydraulic circulation. The primary reaction zone, in which raw water is mixed with chemicals and with a controlled volume of recirculated dense solids, is smaller than in customary solids-contact units. The reaction zone is followed by a flocculation zone or secondary reaction zone where the mixing is less intense and where coagulants or coagulant aids may be added. These two zones are followed sequentially by a third zone where the major portion of the solids separate from the treated water. At the bottom of the basin the high-density solids remain fluidized under semiquiescent condition. It is from this zone that the dense solids are recirculated into the reaction zone.

TYPICAL DENSATOR OPERATING RESULTS

Plant A — Surface Water: 77°F

	Raw Water	Densator Effluent *
Calcium hardness (ppm as $CaCO_3$):	120	80
Magnesium hardness (ppm as $CaCO_3$):	68	27
M Alkalinity (ppm as $CaCO_3$):	126	44
P Alkalinity (ppm as $CaCO_3$):	0	32
Solids in sludge blowdown (grams/liter):		60

Plant B — Well Water: 46°F

	Raw Water	Densator Effluent *
Calcium hardness (ppm as $CaCO_3$):	128	22
Magnesium hardness (ppm as $CaCO_3$):	83	26
M Alkalinity (ppm as $CaCO_3$):	264	124
P Alkalinity (ppm as $CaCO_3$):	0	76
Solids in sludge blowdown (grams/liter):		75

Plant C — Well Water: 53°F

	Raw Water	Densator Effluent *
Calcium hardness (ppm as $CaCO_3$):	248	86
Magnesium hardness (ppm as $CaCO_3$):	156	66
M Alkalinity (ppm as $CaCO_3$):	308	66
P Alkalinity (ppm as $CaCO_3$):	0	44
Solids in sludge blowdown (grams/liter):		170

* Filtered prior to analysis

Plant D — Combination Well and Surface Water: 80°F

	Raw Water	Densator Effluent*
Calcium hardness (ppm as $CaCO_3$):	114	26
Magnesium hardness (ppm as $CaCO_3$):	104	22
M Alkalinity (ppm as $CaCO_3$):	208	45
P Alkalinity (ppm as $CaCO_3$):	0	29
SiO_2 (ppm as SiO_2):	24	7
Solids in sludge blowdown (grams/liter):		120

Plant E — Final clarifier effluent from municipal sewage activated sludge plant

	Influent	Densator Effluent*
Total hardness	200	100
P Alkalinity (ppm as $CaCO_3$):	0	100
M Alkalinity (ppm as $CaCO_3$):	250	150
Total phosphate (ppm as PO_4):	30	0.7
COD (ppm):	60	50
SiO_2 (ppm as SiO_2):	35	25

Infilco DENSATOR plant (center of photo) and Infilco ACCELATOR plant (foreground) soften water at Motorola Inc Phoenix, Arizona
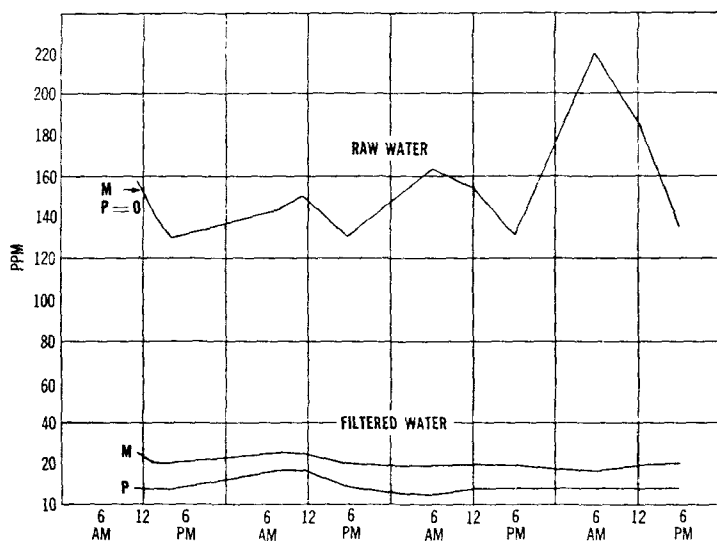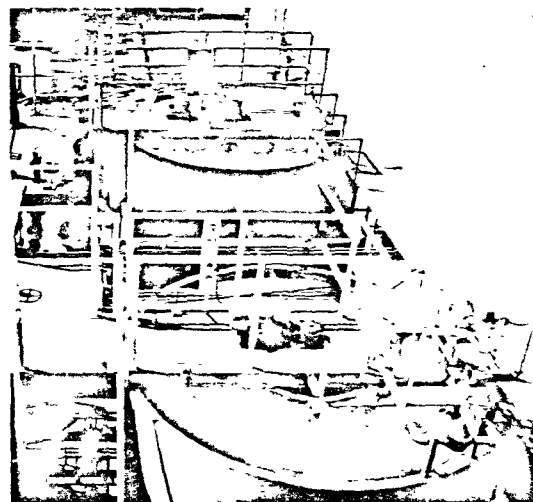
Operating results from a Densator with automatic conductivity ratio control of lime feed

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GATX

PHOSPHORUS REMOVAL BY HIGH-DENSITY, SOLIDS-CONTACT TERTIARY TREATMENT

by

C. F. Garland*

It is conceivable that today's wastewater treatment processes may eventually yield under the combined pressures of increasingly stringent effluent quality requirements and developing technology to systems yet unknown. It appears, however, that in the foreseeable future advanced treatment will largely entail addition of physical-chemical separation processes to the secondary systems now used (1).

Processes of this type for phosphorus removal comprise application of precipitation and/or adsorption reactions induced by addition of suitable cations to the treated wastewater under proper conditions, followed by clarification for suspended solids removal. Practical possibilities in this regard are largely limited to use of lime and iron or aluminum salts, singly or in combination. Numerous studies have shown that any of these treatments is effective at relatively high dosages. Typical data for treatment of effluent from a heavily loaded sewage activated sludge system are presented in Fig. 1(2).

Whereas similar phosphorus reductions are obtained with any of these chemicals, there are other important considerations. It has been observed that lime treatment of high-alkalinity wastewaters produces granular solids which settle, thicken, and dewater readily. Solids from treatment of low-alkalinity waters are more flocculent. Disposal of the waste sludge and recovery of usable lime by calcination is practical.

Further comparison of the treatment possibilities reveals serious relative disadvantages for the iron and aluminum processes. Chemical costs are high and the solids produced are light and voluminous. Treatment with either chemical leads to sludge handling and disposal problems which in a practical sense are as yet unresolved. All things considered, the lime process is favored currently.

High-density, Solids-contact Treatment

Reaction kinetics favor use of a solids-contact process. Several years ago, INFILCO initiated a program to develop a new process of this type with the objectives of lower capital cost, maximum results, and minimum waste sludge volume. The high-density, solids-contact process is the result of that program, and the DENSATOR unit is the device in which the process is applied.

Improved overall treatment results are obtained if the solids concentration in the reaction zone is maintained much higher than is possible in conventional solids-contact units. In most such units the precipitates from 5 to 10 reactions are carried. The DENSATOR unit utilizes the precipitates from up to 50 or more reactions. Other important advantages accruing from this type of treatment are improved solids settling characteristics and a significant reduction in waste sludge volume as a result of the solids densification which occurs after many reactions.

High-density, solids-contact treatment requires special modification of the usual solids-contact equipment. Fig. 2 illustrates some of the features of the DENSATOR unit. The primary reaction zone - in which the influent wastewater, lime, and a controlled volume of recirculated dense solids are mixed - is smaller than in customary solids-contact units. The primary reaction zone is followed by a flocculation or

secondary reaction zone where mixing is less intense and where coagulants or coagulant aids may be added. These two zones are followed sequentially by a third zone in which solids separation occurs. Fluidized high-density solids are accumulated in the bottom of the unit and it is from here that solids are recirculated to the primary reaction zone.

Process Application and Performance

Ten pilot-plant investigations of renovation of treated municipal sewage involving INFILCO solids-contact units on a 7,000 to 35,000-gpd scale have been completed or are in progress in this country and overseas. With the exception of two studies on oxidation pond effluent, this work was on treated waste from various types of activated sludge plants. In addition, a 500,000-gpd DENSATOR unit is being furnished to the Water Reclamation Research Center at Dallas, Texas, for study of chemical treatment of domestic sewage from any of five sources within the White Rock 2-stage trickling filter plant, plus effluents from pilot-scale activated sludge systems. Objectives of these investigations are variable, but all of this experience is pertinent to phosphorus removal.

Phosphorus removal by the lime process improves with increasing pH. This is well illustrated in Fig. 3 based on performance data obtained by Los Angeles County Sanitation Districts at Pomona, California, for single-stage DENSATOR treatment of activated sludge effluent. Influent total phosphorus averaged close to 11 mg/l as P and total alkalinity and hardness 267 and 204 mg/l, respectively, during this 10-week study. In this case, it was necessary to raise the pH to about 10.0 to achieve 80 per cent removal of phosphorus. The hydrated lime dosage for this level of removal was approximately 280 mg/l. Higher dosages increased phosphorus removal to more than 90 per cent. Waste-sludge volume during this study ranged 0.1-0.5 per cent of the throughput

With the possible exception of small plants, the pH of lime-
treated effluents may be too high for direct discharge. Adjustment of
pH, probably by recarbonation, will be necessary when the buffer capa-
city of the receiving stream is insufficient.

The lime demand of treated sewage varies with changes in the
sewage composition and is of course influenced by the alkalinity, being
relatively high when alkalinity is high and lower when the alkalinity is
less. The quantity required to attain a given pH can be quite variable,
as the Pomona data of Fig. 4 show. Because phosphorus removal is pH-
dependent, this is important in terms of process control. Flow-propor-
tioned dosage control leaves something to be desired and pH-controlled
lime feed involves undesirable control problems, cost and maintenance.

In this connection, the double reaction-zone design of the DENSA-
TOR unit enables split treatment and use of a simple and inexpensive auto-
matic ratio control for lime feed when the influent wastewater contains
sufficient alkalinity and calcium hardness. In this process (Fig. 5) a
portion of the influent is introduced into the primary reaction zone with
a high concentration of recycled solids for rapid completion of reactions
at high pH. Excess hydrate alkalinity in proportion to influent alkalinity
is maintained in this zone by lime fed under conductivity ratio control
and "recarbonation" is accomplished in the secondary reaction zone with the
remaining fraction of influent flow. When a DENSATOR unit is coupled with
automatic ratio control of lime feed, the lime system can be greatly
simplified.

The effectiveness of each of these control systems is under study
at the FWPCA-DC demonstration facility at the Blue Plains treatment plant
in Washington, D.C.

As the costs of new water supply and wastewater treatment for disposal coverge, reuse potentials for the treated wastewater should not be overlooked. The wastewater effluent literally becomes too valuable to discard when the cost of up-grading its quality to meet market requirements becomes competitive with the cost of alternative sources of water supply.

Hardness and alkalinity reductions are enhanced only by extensive recycle of sludge and no amount of sludge recycle practically attainable appears sufficient to obtain the degree of hardness and alkalinity reductions experienced with natural waters. Excellent reduction of alkalinity can be obtained by conducting the lime treatment in two stages. In the first stage, an excess of lime is added and the sludge consisting of precipitated calcium carbonate, calcium phosphate, and particulate organic matter is allowed to separate. The clarified water is recarbonated in the second stage in the presence of recycled calcium carbonate sludge. This process is being evaluated at Blue Plains.

Lime-treated effluents can be polished by filtration for further substantial reduction of phosphorus, as well as BOD and suspended solids. Performance of a sand filter following the DENSATOR unit at Pomona is shown in Fig. 6. At pH 10.0, effluent phosphorus was down to 0.25 mg/l, an overall removal of better than 97 per cent.

Additional pilot-plant data from a lengthy investigation at Tucson, Arizona, are presented in Table 1. Recarbonation ahead of filtration is most likely indicated for full-scale installations of this type.

TABLE 1

Performance of Pilot-scale, Lime-process DENSATOR
and Sand Filter on Activated Sludge Effluent at
Tucson, Arizona

Constituent, mg/l	Influent	Sand Filter Effluent
pH	7.6	10.5
Total phosphorus, as P	10	0.23
Chemical oxygen demand	60	50
Phenolphthalein alkalinity, as $CaCO_3$	0	100
Total alkalinity, as $CaCO_3$	250	150
Total Hardness, as $CaCO_3$	200	100
Turbidity, SJU	10	2
Ammonia nitrogen, as NH_3	25	25
Silica, as SiO_2	35	25

Split treatment was used in the Tucson DENSATOR. Chemical treat-ment comprised 310 mg/l of hydrated lime fed as a paste under conductivit ratio control plus 20 mg/l of ferrous sulfate coagulant. A minimum slurr concentration in the primary reaction zone of four per cent by weight was achieved by sludge recycle during this study. Sludge blowdown was only 0.13 per cent of the throughput by volume and contained 30 per cent solid; by weight.

Cost Considerations

The estimated capital cost of a 10-mgd, single-stage, lime-process plant for phosphorus removal is $425,000. For debt service at 4.5 per cent for 25 years, the unit cost of capital amortization is 0.8 cents/1000 gallons of wastewater treated. Further assumptions of chemical use as 300 mg/l of hydrated lime plus 50 mg/l of ferrous sulfate coagulant and sludge disposal by a 25-mile haul to land fill lead to operating and maintenance costs of 3.5 cents/1000 gallons. The total cost of 4.3 cents/1000 gallons is viewed as a conservative figure which would reduce to about 3.5 cents/ 1000 gallons if sludge disposal and lime recovery by calcination were

adopted (3)(4). Inasmuch as the lime dose necessary will depend on the wastewater alkalinity and the degree of phosphorus removal desired, actual costs will differ somewhat from these hypothetical figures in response to specific requirements and conditions.

Recarbonation, when required, adds one cent to the above costs. Two-stage lime treatment on this same scale, including on-site recalcination, will cost approximately 6.5 cents/1000 gallons, and the estimated total unit cost for filtration only is 3.5 cents/1000 gallons.

The cost of treatment in this manner beyond that provided by secondary systems as now designed must of course be added to present costs. Accordingly, appraisal of current practice in the light of new requirements such as phosphorus removal is warranted. Through a contract research program with FWPCA's Advanced Waste Treatment Group at Cincinnati, INFILCO has examined the potential for an integrated sequential combination of biological and chemical treatment (2).

Good biological removal of soluble organics can be realized in much less time than is presently the case in practice if the requirement for bioflocculation in activated sludge treatment can be overcome by appropriate subsequent chemical treatment. From laboratory and pilot-plant study of a highly-loaded, short-detention activated sludge process followed by a solids-contact chemical treatment system it was concluded that the combination system is capable of exceeding performance of conventionally designed activated sludge systems and offers important advantages in treatment stability and flexibility. The greater operating cost of the two-stage system is offset by lower capital cost and it appears that such plants can be constructed and operated at the same total cost as that of the traditional system.

Lime flocculation of Tucson sewage at pH 8.9 in the second stage lowered the influent total phosphorus concentration of 9.0 down to 0.76 mg/l, a removal of over 90 per cent. Thus, particularly when removal of phosphorus is a treatment objective, a system of this type has applicatic to modification or expansion of existing activated sludge installations and new-plant construction.

Additional INFILCO research reported two years ago in a series of national seminars entails use of a DENSATOR-filter combination as the firs component of a complete wastewater renovation system (5). Following reduction of suspended solids, phosphorus, hardness and alkalinity in the pretreatment system, the process continues with adsorption of organic matter by thermally regenerated activated carbon and demineralization by multi-bed ion-exchange. The product water from such a system of course contains no phosphorus and its quality in other respects is excellent. Pro·duct recovery for the overall system exceeds 90 per cent.

On a 10-mgd scale, the high-quality water will cost about 40 cents/ 1000 gallon (the exact figure is strongly influenced by the dissolved-solids content of the local water supply.) This cost, although high, compares favorably with figures for alternative complete wastewater systems under study. Where water of intermediate quality is suitable, the concepts of split treatment, partial demineralization, and blending may be applied to advantage and can generate significant economy.

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3. Smith, Robert, "Cost of Removing Phosphorus From Wastewater," FWPCA Seminar on Phosphorus Removal, Chicago, Illinois (May 1-2, 1968).

4. Smith, Robert, "A Compilation of Cost Information for Conventional and Advanced Wastewater Treatment Plants and Processes," FWPCA Cincinnati Water Research Laboratory (December, 1967).

5. "INFILCO National Seminars," a report published by INFILCO/Fuller Company, Tucson, Arizona (1966).
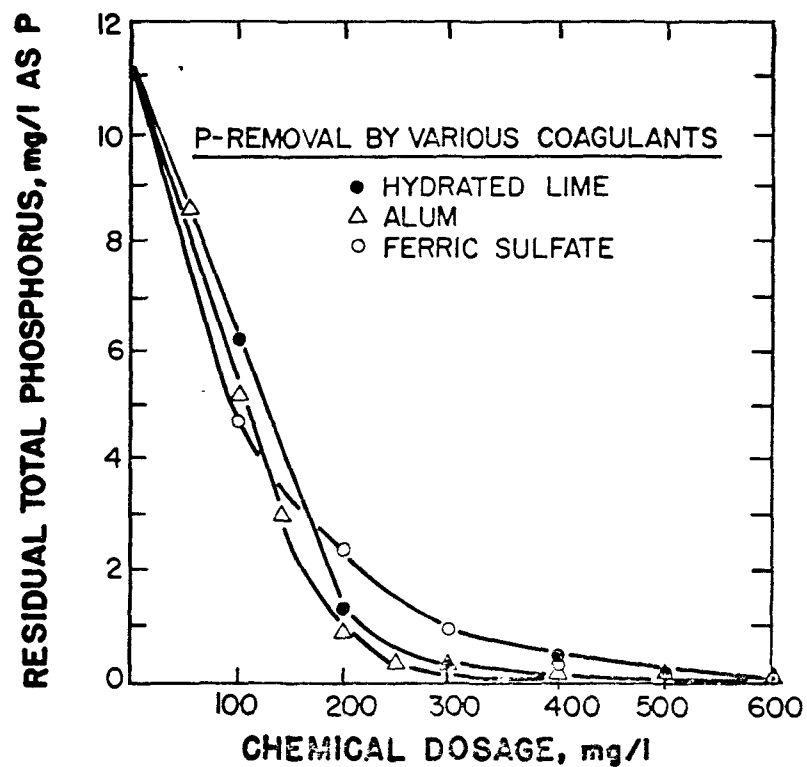
Figure 1 - Phosphorus removal by various chemicals. Labora-
tory data for treatment of effluent from a heavily
loaded activated sludge system.
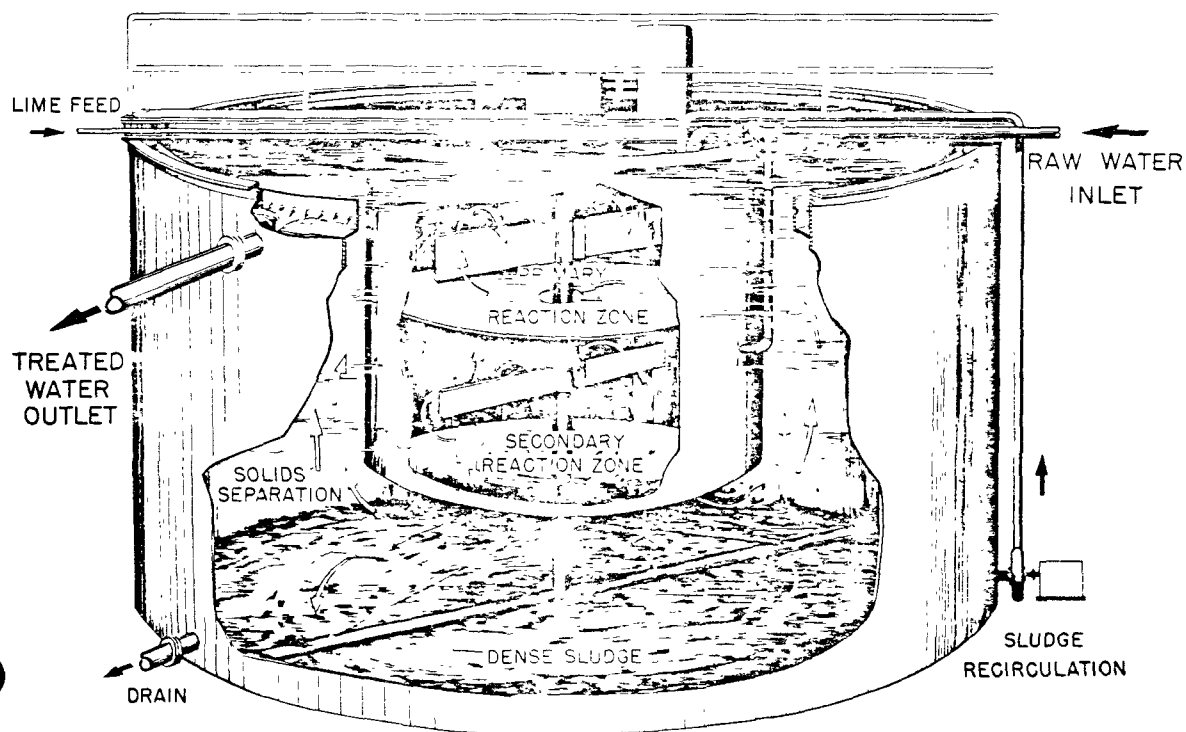
Figure 2 - The DENSATOR high-density, solids-contact unit.
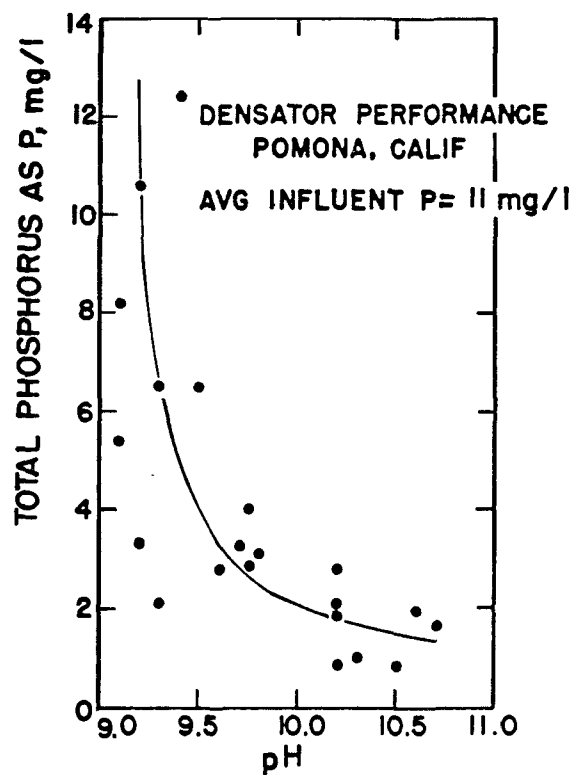
Figure 3 - The effect of pH on phosphorus removal by DENSATOR
lime treatment of activated-sludge effluent at
Pomona, California (LACSD data).
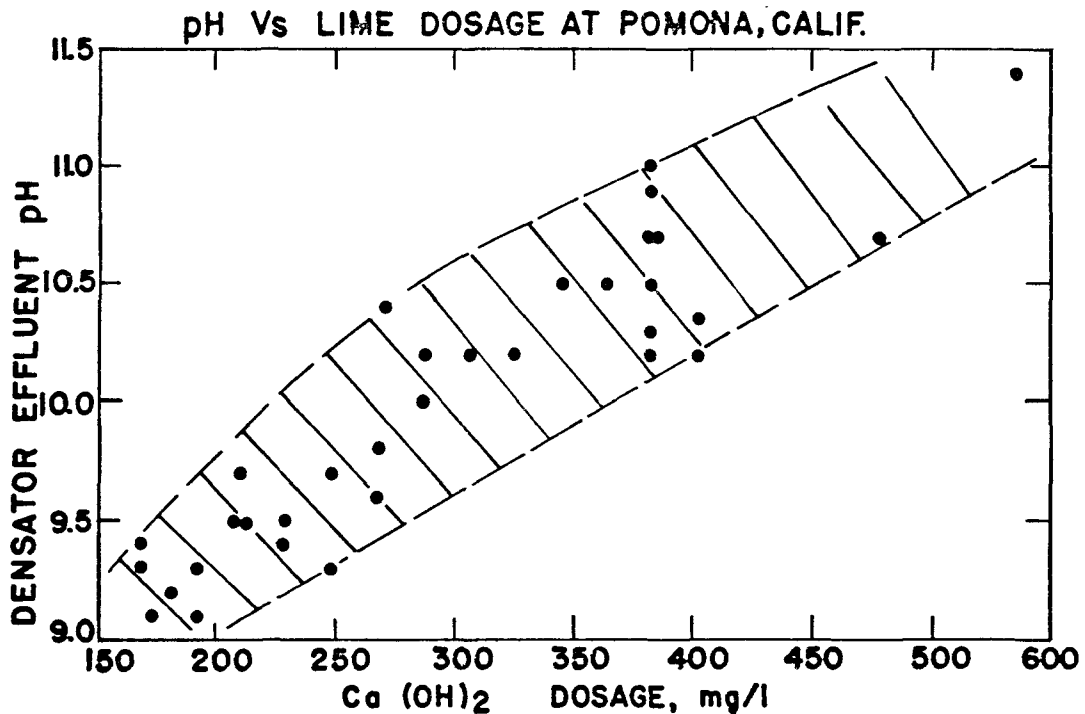
Figure 4 - The relationship between lime dosage and pH for
DENSATOR lime treatment of activated-sludge
effluent at Pomona, California (LACSD data).
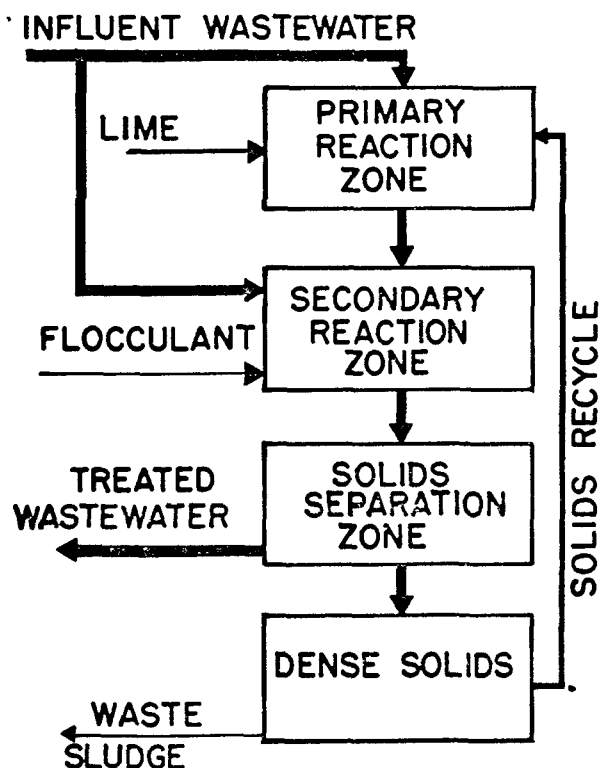
Figure 5 - Flow diagram for split treatment in the DENSATOR
high-density, solids-contact unit.
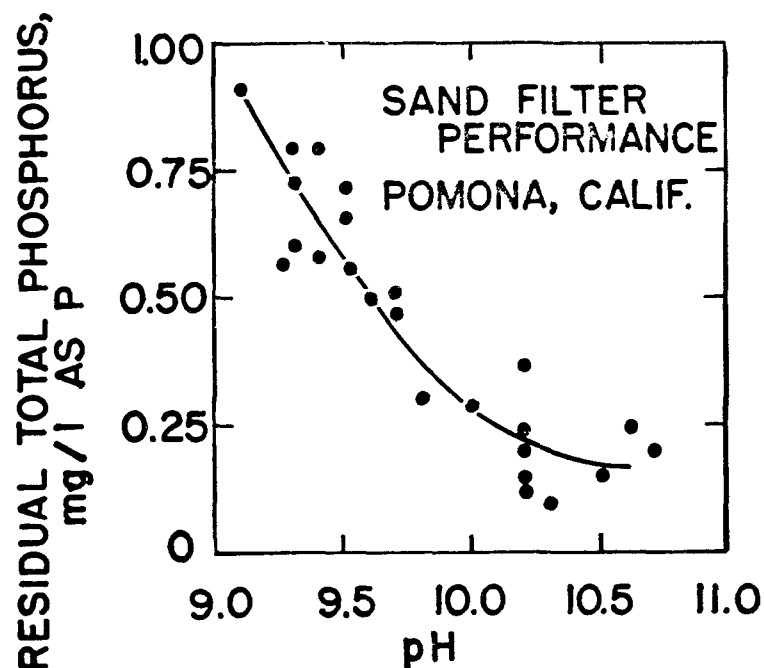
Figure 6 - Residual phosphorus versus pH for lime-treated
and filtered activated sludge effluent at
Pomona, California

360° PERIPHERAL LAUNDER

"D" RAW WATER INFLUENT

WALKWAY

"A"

SLUDGE RECYCLE

"E" EFFLUENT

PLAN

"C" DRIVE

"H" PERIPHERAL LAUNDER

SLUDGE RECYCLE

EFFLUENT

SECONDARY WATER FLOW

"B"

"F"

"G"

SAMPLE COCKS

MANHOLE

SECTIONAL ELEVATION

CAPACITY	GPM	
A		INSIDE DIAMETER
B		DEPTH
C		DRIVE (H P)
D		INFLUENT
E		EFFLUENT
F		OVERFLOW
G		SLUDGE BLOW-OFF AND DRAIN
H		" WIDE x "DEEP LAUNDER

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INFILCO DENSATOR TREATING PLANT

Steel Side Sheet with
Concrete Bottom
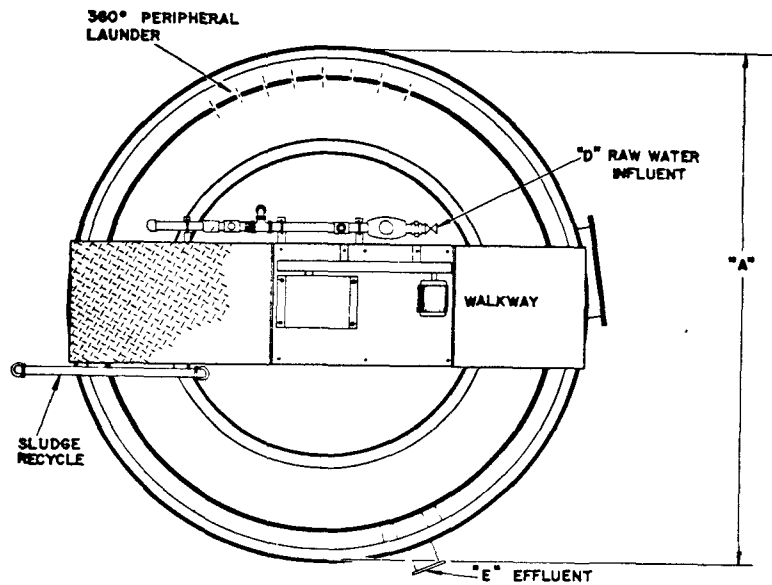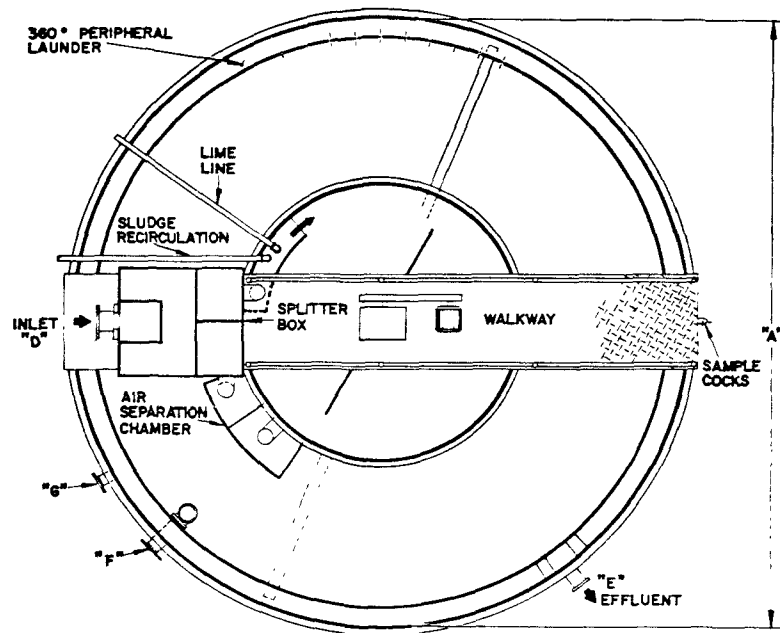
PLAN
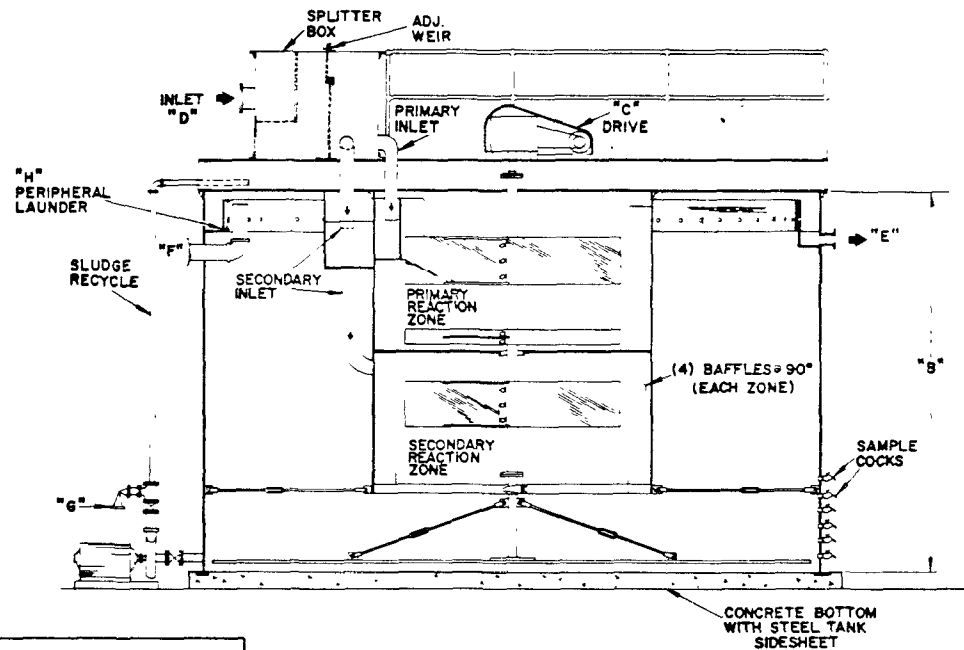
SECTIONAL ELEVATION

CAPACITY	GPM
A	INSIDE DIAMETER
B	DEPTH
C	DRIVE (HP)
D	INFLUENT
E	EFFLUENT
F	OVERFLOW
G	SLUDGE BLOW-OFF AND DRAIN
H	" WIDE x " DEEP LAUNDER

FULLER COMPANY / GENERAL AMERICAN TRANSPORTATION CORPORATION
INFILCO PRODUCTS

GATX

INFILCO DENSATOR TREATING PLANT

Steel Side Sheet with
Sloping Concrete Bottom

360° PERIPHERAL LAUNDER

LIME LINE

SLUDGE RECIRCULATION

INLET "D"

SPLITTER BOX

AIR SEPARATION CHAMBER

WALKWAY

SAMPLE COCKS

"A"

"G"

"F"

"E" EFFLUENT

PLAN

SPLITTER BOX

ADJ. WEIR

"C" ROTOR DRIVE

"J" SCRAPER DRIVE

PRIMARY INLET

INLET "D"

"H" PERIPHERAL LAUNDER

"F"

PRIMARY REACTION ZONE

"E"

SLUDGE RECYCLE

SECONDARY INLET

(4) BAFFLES ∓90° (EACH ZONE)

"B"

SECONDARY REACTION ZONE

SAMPLE COCKS

"G"

SCRAPER

CONCRETE BOTTOM WITH STEEL TANK SIDESHEET
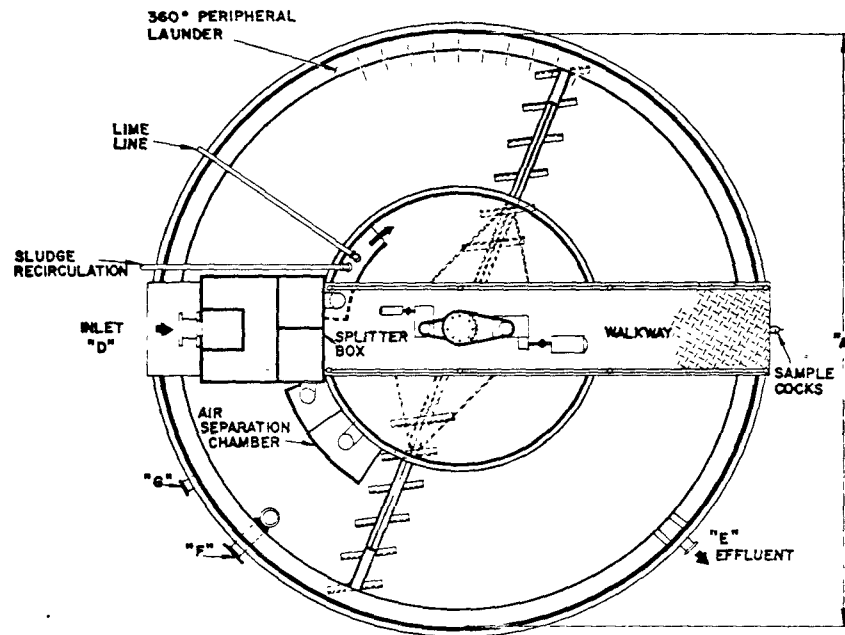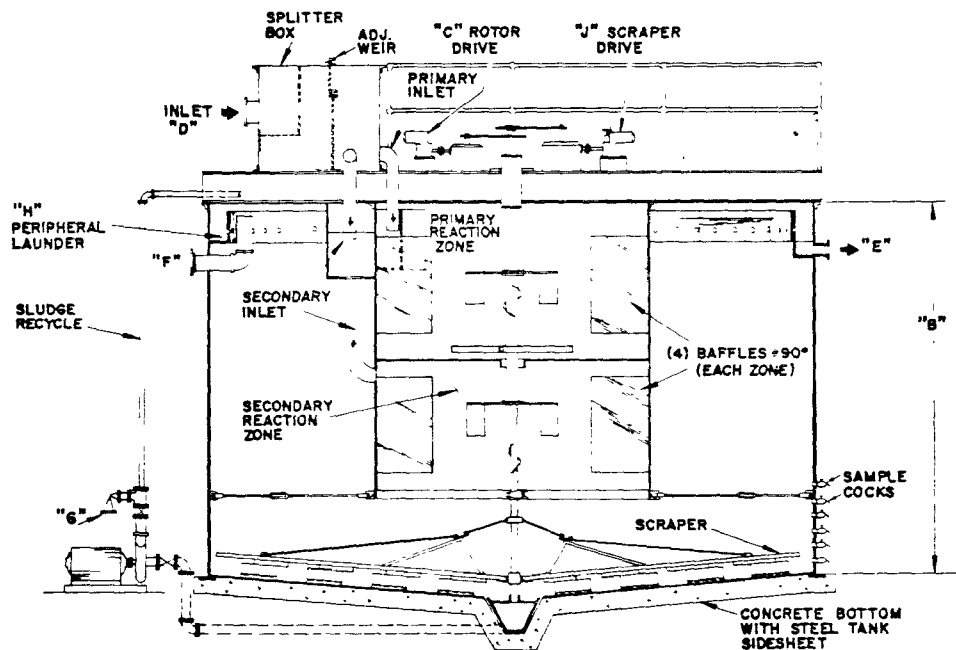
SECTIONAL ELEVATION

CAPACITY	GPM
A	INSIDE DIAMETER
B	DEPTH
C	ROTOR DRIVE (HP)
D	INFLUENT
E	EFFLUENT
F	OVERFLOW
G	SLUDGE BLOW-OFF AND DRAIN
H	" WIDE x "DEEP LAUNDER
J	SCRAPER DRIVE (HP)

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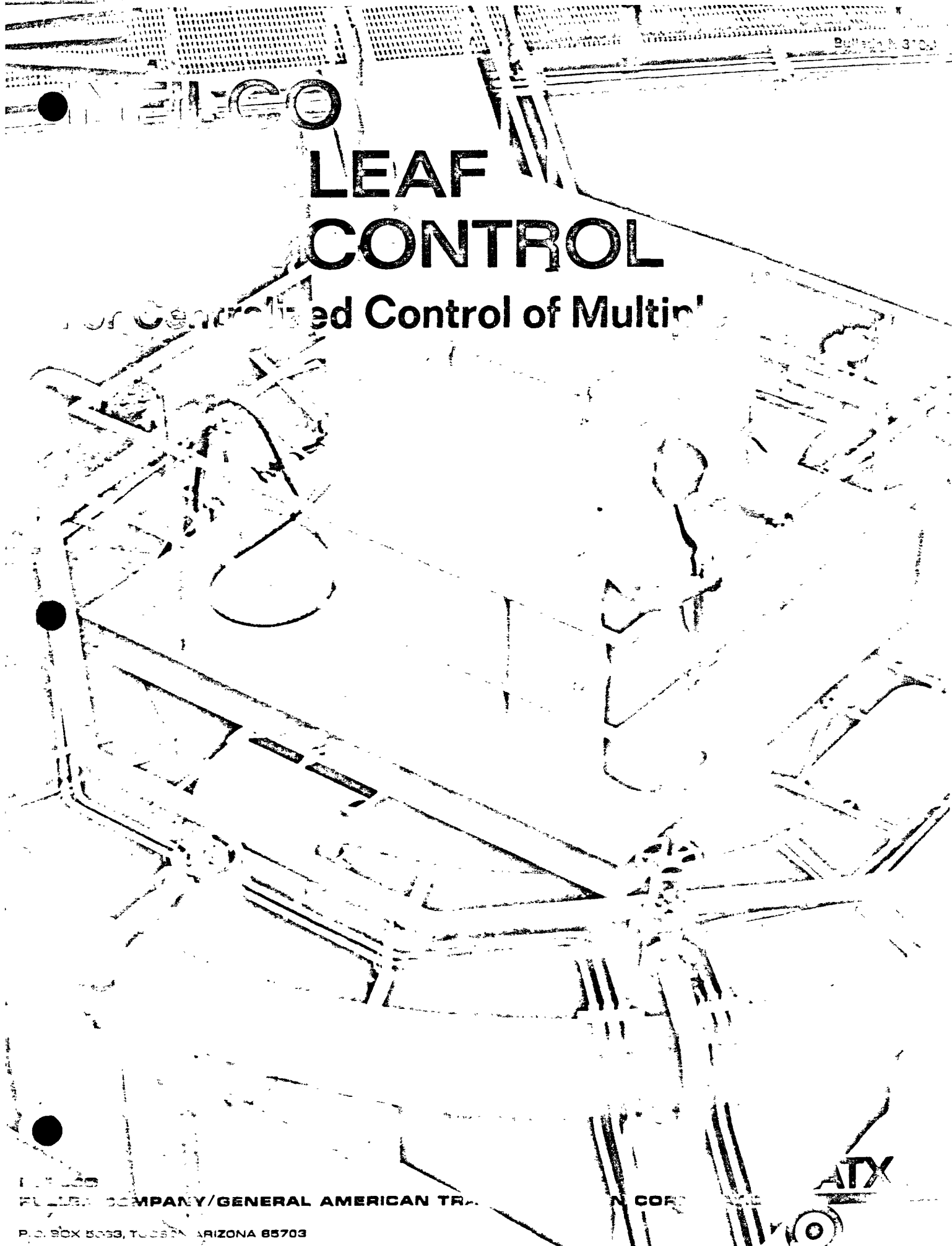
LEAF
CONTROL

For Centralized Control of Multiple
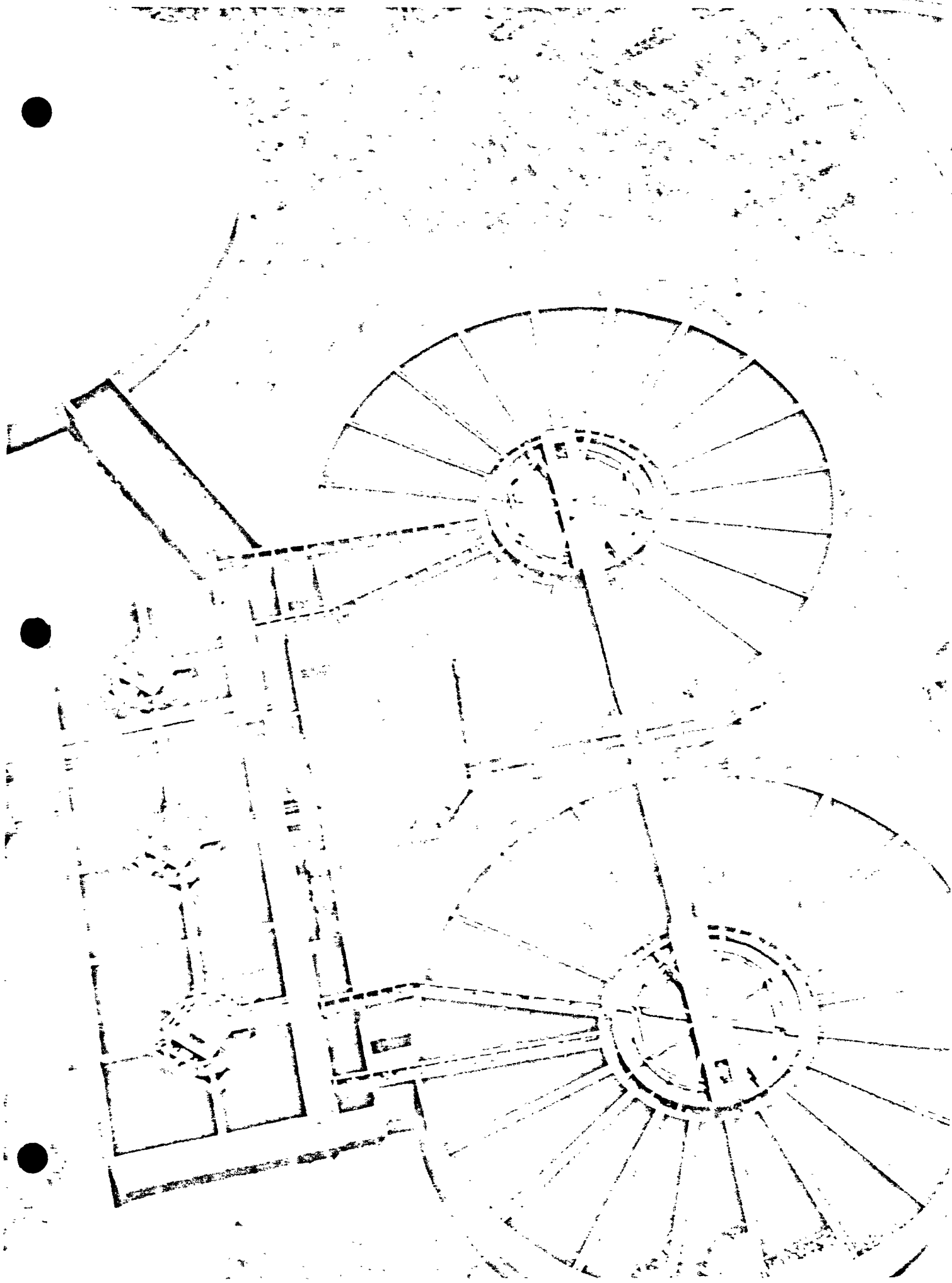
GREENLEAF FILTER CONTROL

○ **Improved filter plant performance**

○ **Lower installed cost**

○ **Minimum space requirements**

○ **Reduced maintenance costs**

Flexibility of filter plant design

The Greenleaf Filter Control offers a new design concept for **controlling** the functions of multiple rapid sand gravity filters, utilizing conventional low head loss underdrain systems, filter media, and filtration rates. It is **not** a new filter design. The filter design is by the consulting engineer or industrial water treatment specialist—incorporating his own and regulatory agency standards. By incorporating the Greenleaf Filter Control Center in the design of filter plants, the consulting engineer can offer a superior and highly economical method of control in his filter installation.

ıproved Filter Plant Performance

:ers are always under positive head

s impossible to apply a negative head or draw air into
ə filter bed because the filtered water discharges over
 adjustable effluent weir above the elevation of the
:er media.

possible to shock or surge a filter

e Greenleaf Filter Control provides an **ideal filter cycle,**
ıce all rate of flow increases and decreases are gradual.
is impossible to upset the filter bed by sudden in-
eases in backwash flow or by pulsating flows due to
alfunction of controllers during operation.

ɔw is equally and automatically divided among the ters● peration

ɔ mechanical flow controllers are required—flow control
nong the filters is achieved by basic hydraulics. Rec-
ngular weirs with free fall at each filter cell equally
vide the inflow. When one filter cell is being back-
ashed, its inlet siphon valve is stopped and total flow
 equally and simultaneously divided to the operating
ter cells.

mplicity of operation

II filter control functions are located in a control center
round which the filter cells are placed. The operator
ın easily select and control any individual filter cell from
.e common control location. All controls are visible
ɔove the filters. Accessibility, convenience of a com-
on location, and choice of local or remote operation are
rovided—any degree of automation can be included.
entralized control means simple, economical operation.

ackwash water can be provided by the filtering cells

ilters and control centers are designed to provide the
lter rates specified. Multiple filter control units (two or
nore filter control centers) allow one filter cell to be
ack●hed from filtered effluent of the other filter
ells without drawing on clearwell capacity.

iiphon valves contribute to efficient operation

`he siphon valves control the filtering and backwashing
ycles and are extremely simple in operation. These
valves are connected to a vacuum system by small indi-
vidual 3-way actuating valves to initiate and stop the flow
of water during filtering and backwashing.

Reduces Both Initial and Installation Costs

Careful evaluation of all major equipment and installa-
tion costs indicate a savings of approximately 20% in
the cost of a 5 mgd filter plant. In larger plants, greater
savings can be expected.

The Greenleaf Filter Control eliminates

Filter pipe galleries · interconnecting filter piping ·
line-size filter operating valves · filter effluent control-
lers and control systems · filter gauges · backwash
controllers and gauges · pneumatic control systems.

Pre-fabrication reduces field installation costs, insures design accuracy

The ease and economy of installing the Greenleaf Filter
Control significantly reduce field construction costs. The
central control center is pre-fabricated and shipped in a
minimum number of sub-assemblies. Pre-fabrication of
the control center assures design accuracy.

The central control compartment which includes the
backwash siphon valves, inlet weir chamber controls and
inlet feed channels is in one pre-assembled unit. (The
D-4 unit is shipped in quadrants.)

The control compartment is placed on a flat concrete
slab and grouted in place. The annular feed channel and
inlet siphon valves are then assembled onto the control
compartment.

Excavation problems minimized in clearwell construction

Because of the positive head design of the Greenleaf
Filter Control, no deep excavations for the clearwell are
required since the clearwell need not be located under
the filters as in conventional designs. This results in a
great savings under conditions of improper soil or rock
foundation. Additionally, the clearwell site can be selec-
ted for optimum land usage.

With imposing any design restrictions on the individual filter plant, the Greenleaf Filter Control offers a functionally perfect and simply operated control of the filter plant. It has substantial cost, space and maintenance advantages.

Lower Maintenance Costs

The Greenleaf Filter Control assures reduced maintenance costs:

Visible controls

All controls are visible above the filters and readily accessible for maintenance and servicing.

Design simplicity

Simplicity of design and minimum use of equipment with moving parts substantially reduce maintenance requirements.

Minimized components

Elimination of large valves, extensive piping runs, pneumatic flow controls and instrumentation significantly reduce maintenance costs.

Flexibility of Filter Plant Design

The Greenleaf Filter Control allows designs for flows of less than 1 mgd up to large multi-mgd systems. A single filter control center can be used for capacities up to approximately 9 mgd; multiple control centers are used for higher flows. Filter construction may be round, square or rectangular depending on design criteria.

Minimum Space Requirements

We have computed by actual comparison a space saving of approximately one-third of that required for a like-capacity plant utilizing conventional filter control systems. The arrangement of the filter bays around the control center eliminates most of the connecting and control equipment of conventional filter controls, and provides an extremely compact filter plant. Additional space saving is realized by the use of siphon valves for the filtering and back-washing cycles to replace the conventional line-size water valves. The drawing below demonstrates the relative space requirements of like capacity filter plants using the Greenleaf Filter Control and conventional controls.
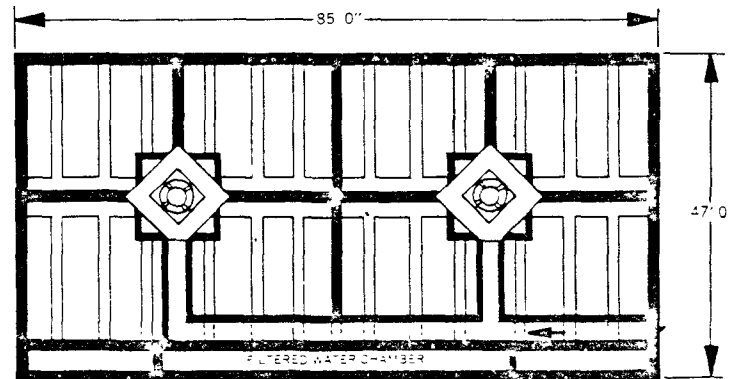
FILTER PLANT WITH GREENLEAF FILTER CONTROLS
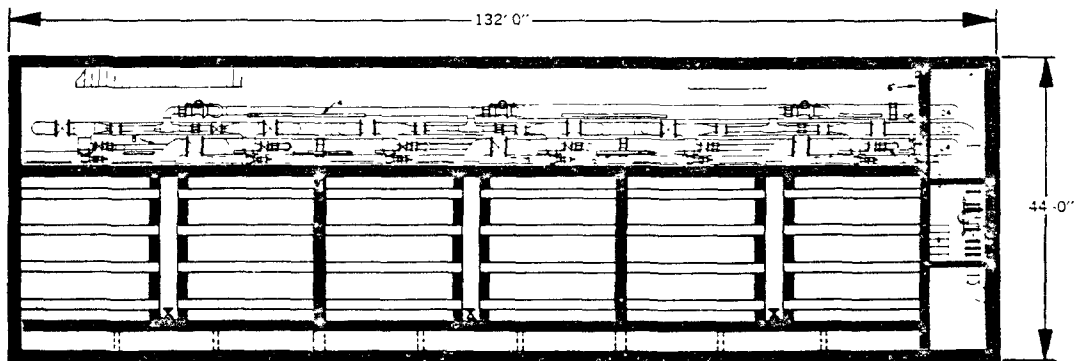
FILTER PLANT WITH CONVENTIONAL CONTROLS

Relative Elevations

All of the following cross-sectional drawings of filter cells and Greenleaf controls illustrate the functions of opera-tion. In potable water installations, double wall con-struction at certain locations and other special features may be included as required.

The elevation of the filtered water effluent control weir is the datum point from which available head for both filtering and backwashing is measured. Approximately 3'-0" from the top of the wash trough to the effluent con-trol weir will provide a 15 gpm/sq. ft. wash rate if a lo head loss type of underdrain is used. Adjustment of th elevation of the effluent control weir can change the ava able head and therefore the wash rate. The dimensi from the effluent control weir to the high water level a filter cell is the maximum available head for filterin

Labels in diagram: DRAIN; UNDERDRAIN; COMMON FILTERED WATER CHAMBER; FILTER MEDIA; WASH TROUGH; FOREBAY; FILTER CELL; FILTERED WATER TO STORAGE OR SERVICE; BACK' HEAD; FILTER HEAD; EFFLUENT CONTROL WEIR; FREEBOARD; TOP OF BASIN; INLET SIPHON VALVE; INLET WEIR CHAMBER; TO VACUUM PUMP; VACUUM TANK; FLOW; PRETREATED WATER FLUME

GLOSSARY OF TERMS

1. **Loss of head** is the resistance of the filter bed to passage of water.

2. **Positive head** is the depth of water in the filter basin above the effluent weir elevation in the filtration cycle and the depth of water in the effluent

3. **Negative head** in a conventional filter control system is the depth of water from the underdrain to the water seal on the effluent pipe.

4. A siphon valve is a siphon with a con-nection at the top to a vacuum sourc which can start a flow through th siphon or be vented to stop the flo through the siphon.

5. Clearwell is the filtered water reservoir

6. A weir is a spilling edge used for con-trolling liquid flow.

weir chamber above the wash troughs in backwash cycle.

Greenleaf filter control installation at paper processing plant
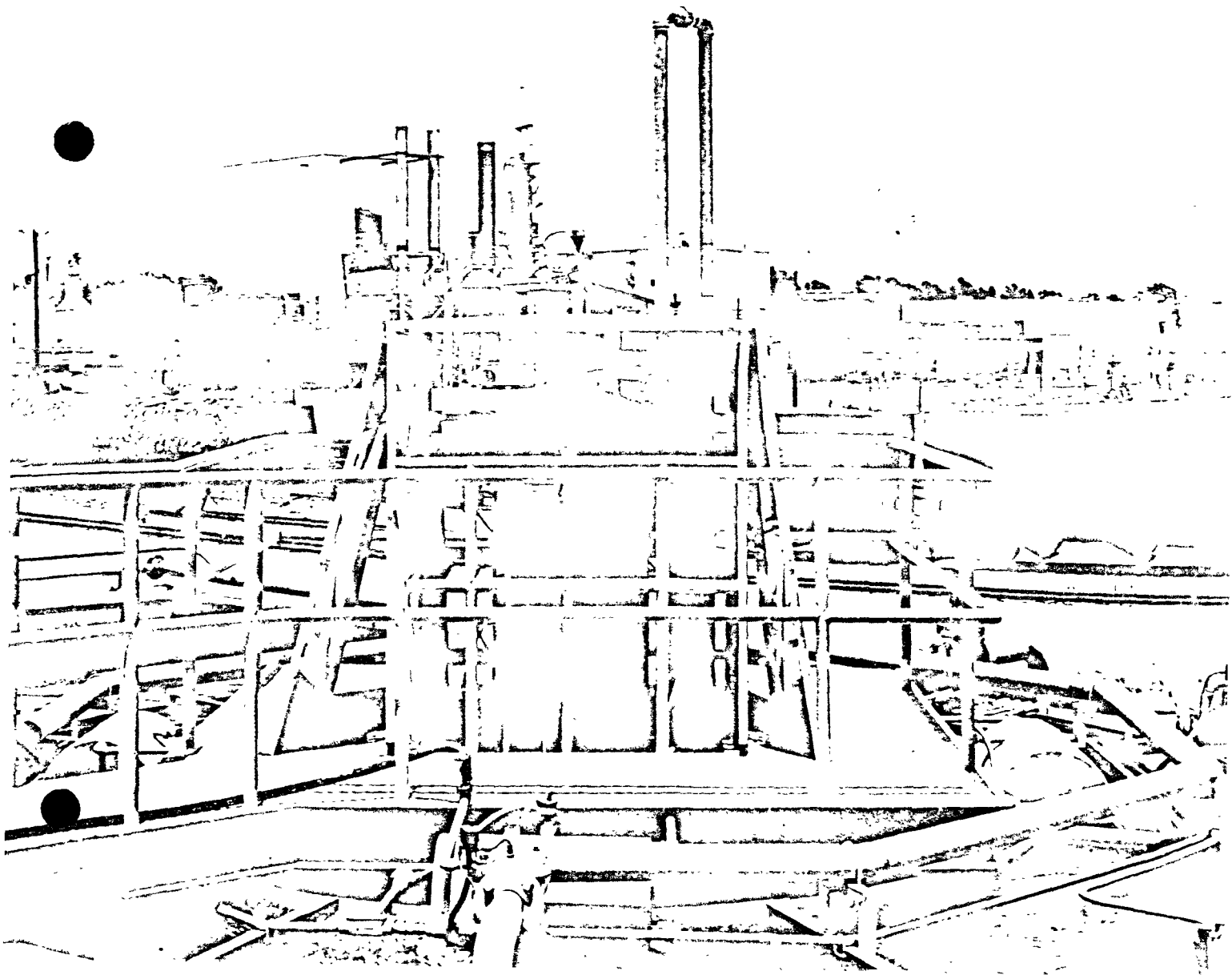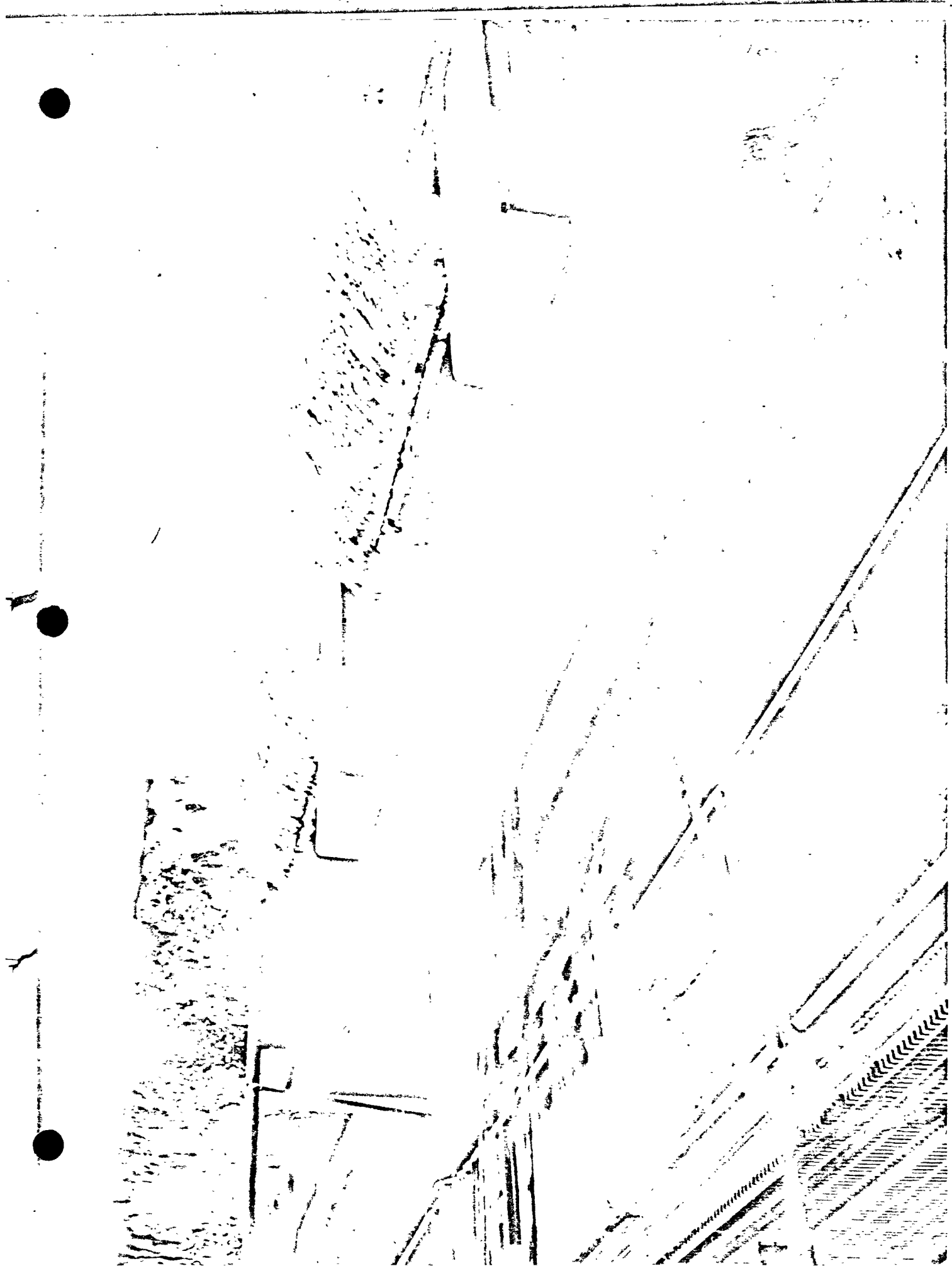
Schematic Filter Cycle

The diagram illustrates the facts that the filter is always under positive head and that all filter functions are completed in a hydraulically perfect cycle. This schematic drawing does not in any way reflect the relative time periods of the filtration and backwash cycles. The zero point on the curve represents the elevation of the effluent control weir. During the filtering cycle, the head on the filter increases above this elevation as the filter unit becomes dirty. After the inlet siphon valve has been stopped, the head or level decreases until hydrostatic balance is achieved. The operation of the backwash siphon valve allows the positive head at the effluent weir to initiate backwashing. As the level in the filter lowers towards the wash troughs, the increasing head builds the wash rate to maximum design which is maintained until the backwash siphon valve is vented. The rate of flow then decreases gradually and allows the bed to settle until the zero point is reached. The inlet siphon valve is then started and the filtering cycle begins anew. All rate increases and decreases are in proportion to the square root of H. A perfect cycle is thereby achieved without the use of rate of flow controls. The operator cannot shock or surge a filter.

VACUUM TANK
TO VACUUM PUMP
(D) INLET WEIR
(C) INLET WEIR CHAMBER
(E) INLET SIPHON VALVE
PRETREATED WATER FLUME
(A)
(A) ANNULAR DISTRIBUTOR CHANNEL
FILTER CELL
FILTER CELL
(E) EFFLUENT CONTROL WEIR
FOREBAY
FOREBAY
WASH TROUGH
FILTER MEDIA
COMMON FILTERED WATER CHAMBER
UNDERDRAIN
DRAIN
BACKWASH WASTE SECTION

START UP

INITIAL FILTRATION OPERATION

VACUUM TANK
PRETREATED WATER FLUME
FILTER CELL
FILTER CELL
(F) EFFLUENT WEIR CHAMBER
(E) EFFLUENT CONTROL WEIR
FOREBAY
FOREBAY
WASH TROUGH
FILTERED WATER TO STORAGE OR SERVICE
FILTER MEDIA
COMMON FILTERED WATER CHAMBER
UNDERDRAIN
DRAIN
BACKWASH WASTE SECTION

10

MID-CYCLE FILTRATION OPERATION

Diagram labels (top figure):
- VACUUM TANK
- PRETREATED WATER FLUME
- TOP OF BASIN
- (H) FILTER CELL
- FOREBAY
- FOREBAY
- FILTER CELL (H)
- (I) ELEVATION OF FILTERED WATER OVERFLOW
- (E) EFFLUENT CONTROL WEIR
- WASH TROUGH
- FILTER MEDIA
- FILTERED WATER TO STORAGE OR SERVICE
- COMMON FILTERED WATER CHAMBER
- COMMON FILTERED WATER CHAMBER
- UNDERDRAIN
- DRAIN
- BACKWASH WASTE SECTION

MAXIMUM LOSS OF HEAD IN ONE CELL

Diagram labels (bottom figure):
- VACUUM TANK
- PRETREATED WATER FLUME
- (L) 3 WAY VALVE (ACTUATES INLET SIPHON VALVE)
- (B) INLET SIPHON VALVE
- (J) BACKWASH SIPHON VALVE
- TOP OF BASIN
- FILTER CELL
- FOREBAY
- FILTER CELL
- (K)
- WASH TROUGH
- FILTER MEDIA
- FILTERED WATER TO STORAGE OR SERVICE
- COMMON FILTERED WATER CHAMBER
- COMMON FILTERED WATER CHAMBER
- UNDERDRAIN
- DRAIN

VACUUM TANK

PRETREATED WATER FLUME

G INLET SIPHON VALVE

FILTER CELL

FILTER CELL

FOREBAY

H

I ELEVATION OF FILTERED WATER OVERFLOW

E EFFLUENT CONTROL WEIR

WASH TROUGH

FILTER MEDIA

FILTERED WATER TO STORAGE OR SERVICE

UNDERDRAIN

COMMON FILTERED WATER CHAMBER

COMMON FILTERED WATER CHAMBER

DRAIN

BACKWASH WASTE SECTION

FILTER CELL DRAIN

INITIATION OF CELL BACKWASH

VACUUM TANK

PRETREATED WATER FLUME

M 3 WAY VALVE (ACTUATES BACKWASH SIPHON VALVE)

J BACKWASH SIPHON VALVE

FILTER CELL

FOREBAY

FILTER CELL

FILTER MEDIA

FILTERED WATER TO STORAGE OR SERVICE

UNDERDRAIN

COMMON FILTERED WATER CHAMBER

COMMON FILTERED WATER CHAMBER

DRAIN

BACKWASH WASTE SECTION

12

VACUUM TANK

PRETREATED WATER FLUME

BACKWASH SIPHON VALVE

FILTER CELL

FILTER CELL

ELEVATION OF FILTERED WATER OVERFLOW

EFFLUENT CONTROL WEIR

FILTER MEDIA

UNDERDRAIN

DRAIN

BACKWASH WASTE SECTION

FILTERED WATER TO STORAGE OR SERVICE

END OF CYCLE (BED SETTLED)

VACUUM TANK

PRETREATED WATER FLUME

3 WAY VALVE (ACTUATES INLET SIPHON VALVE)

3 WAY VALVE (ACTUATES BACKWASH SIPHON VALVE)

BACKWASH SIPHON VALVE

INLET SIPHON VALVE

FILTER CELL

FILTER CELL

EFFLUENT CONTROL WEIR

WASH TROUGH

FILTER MEDIA

UNDERDRAIN

DRAIN

BACKWASH WASTE SECTION

FILTERED WATER TO STORAGE OR SERVICE

Underdrain Systems

Filters designed to utilize the Greenleaf Filter Control should incorporate underdrain systems which operate efficiently with low loss of head during backwash. The Infilco Fre-Flo underdrain (shown below) is recommended in that it fulfills the requirements for low head loss and excellent distribution of backwash water. This underdrain has a life expectancy equal to the rest of the filter plant. The trapezoidal beam type underdrain is also recommended as it reduces installation cost since beams can be cast at the job site during the pouring of concrete for the filter walls.
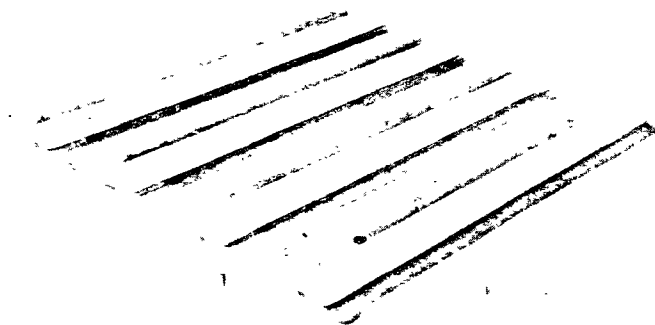
Vacuum Tank

Vacuum is used to actuate the siphon valves. The vol of air to be removed is so small that constant vac is maintained in the tank by minimal operation of vacuum pump. A single tank with a double compartr and drop legs provides the two different vacuum quirements for the inlet and backwash siphon va (approximately 2.0 inches and 15.0 inches of mer respectively) from a single vacuum source. The vac pump is connected to the high vacuum compartr and is controlled by an electrode-type level switc this compartment. The low vacuum compartment is plied from a differential drop leg assembly. The sho low vacuum drop leg is located within the high vac drop leg. The high vacuum leg is open at the bot' but it is sealed by the minimum water level mainta by the drain outlet in the control center. The he of the open bottom of the low vacuum leg above drain elevation is equal to the low vacuum requirem When an inlet siphon valve is actuated, the air in inlet siphon is vented to the low vacuum compartm This air travels to the point of lowest pressure whic the bottom of the low vacuum drop leg, bubbling thro the water in the high vacuum drop leg to the vacuum compartment of the tank. The air enters high vacuum compartment and causes the water to recede, operating the electrode level switch and s ing the vacuum pump to restore the water level t normal position of approximately 17 feet above drain. All points are considered in terms of absc pressure.

14
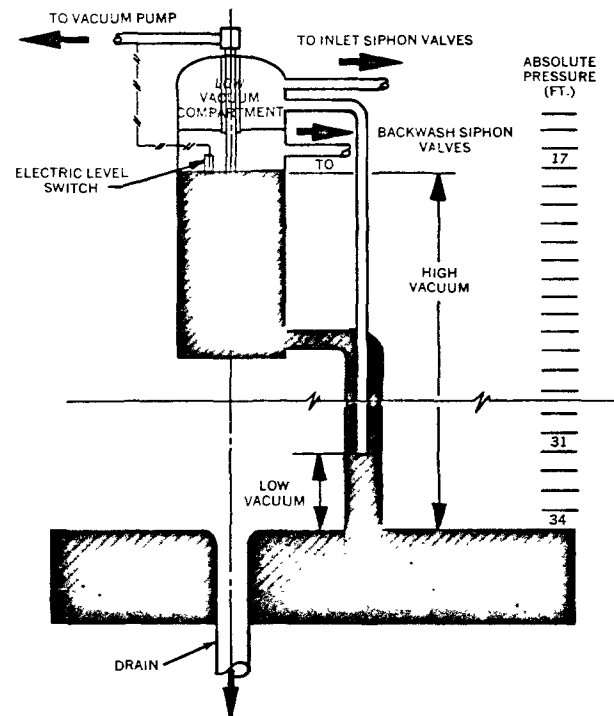
Operating Controls

Operating controls are available in remote manual, semi-automatic or fully automatic designs housed in free standing modular cabinets. With minor electrical changes, any system can be converted to either of the others, or expanded to accommodate additional filter control units.

In the event of power failure, filter cells in service will remain in service and a cell in backwash cycle will return to service. When power is restored. the interrupted backwash cycle will be re-initiated.

The modular cabinets are shipped completely equipped, assembled, wired, piped and pre-tested in the factory.
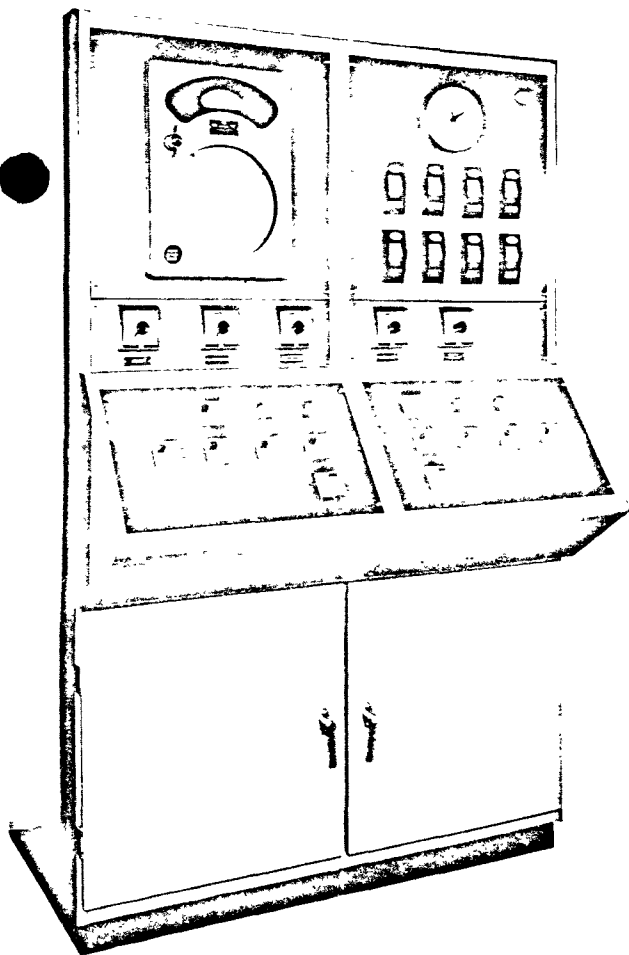
Siphon Valves

The siphon valves of the Greenleaf Filter Control are ON-OFF valves and are **not** flow control valves. Siphon valves are actuated by the 3-way actuating vales at the vacuum tank.

For the filtration cycle, the siphon valves start or stop the flow between the annular inlet flume and the individual compartment of each filter cell. For the backwash cycle, the siphon valves start, maintain or stop the flow between the filter cell forebays and the backwash waste section in the central core. Siphon valves, which have very small head loss, are economical and easy to operate.
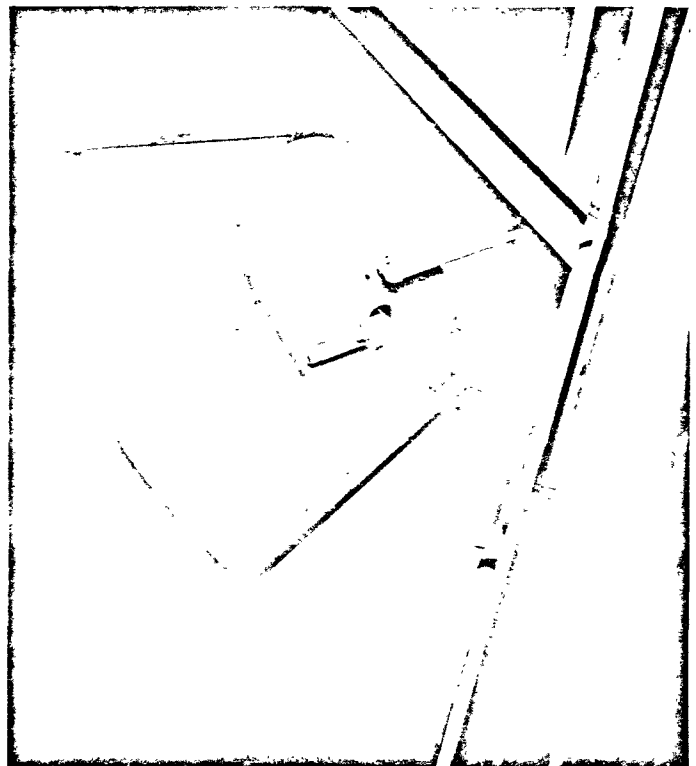
There shall be furnished and installed . . .   . .Greenleaf Filter Control unit(s) having a total capacity
. . . . .gallons per minute when filters are operating at the rate of . . . . .   gallons per mintue per squa
foot. Maximum backwash rate for one filter cell shall be . . . . .gpm.

The Greenleaf Filter Control unit(s) shall be designed to permit a maximum filter head loss of . . . . .fe
The head required for backwashing a filter shall be . . . . .feet. The Greenleaf Filter Control unit(s) sh
be as manufactured by Fuller Company, Tucson, Arizona.

Control Units

Each control unit shall consist of one annual feed cha
nel and individual weir chambers for each of the filt
cells. The water to be filtered shall be introduced to t
annual feed channel through an inlet flume. Inlet siph
valves shall connect the flow of water from annular fe
channel to the weir chambers which regulate the fl
of water over free fall weirs to each filter cell. The bac
wash siphon valves shall be so arranged as to take wat
from each filter cell and deliver it to a central co
partment from which it shall flow to waste over
internal outlet weir.

The central control compartment, including backwa
siphon valves, feed conduit and weir plates shall be
welded construction and shall be fabricated of AISI Ty
304 stainless steel. The annular feed channel and in
siphon valves shall be welded mild steel. The cont
section shall be fabricated in . . . . . .section(s) to
assembled at the project site. The inlet flume shall
of concrete/steel and shall be furnished by the cc
tractor/equipment supplier. A $\frac{3}{16}$" checker-plate ste
walk-way and $1\frac{1}{2}$" pipe handrail shall be provided f
access to the filter control unit.

Outlet Weir

An external adjustable outlet weir for manual cont
of the hydraulic head shall be provided by the equ
ment manufacturer. The weir shall be installed in t
external weir chamber by others.

Vacuum System

A vacuum system for each control center shall be furnished and installed. It shall consist of a two compartment vacuum tank, drop legs, piping connections and 3-way valves connected to the siphon valves, and a sight glass. The 3-way valves shall actuate the siphon valves at the direction of the operator. Piping shall be of the size shown and shall be installed as indicated on the drawings.

Vacuum Pumps

. . . . electric motor driven, single stage, water sealed, rotary type vacuum pumps shall be furnished and installed as shown on the drawings. Each vacuum pump shall be an approved electric motor-driven type, having capacity of   . of air at a vacuum of 15 inches hg. In multiple control center installations, the vacuum pumps shall be manifold to the various vacuum tanks to serve all control centers.

Motors shall be . . . . horsepower, drip-proof/T.E. induction motor suitable for operation on . . . . phase, ... cycle, . . . volt alternating current. There shall be furnished and installed in the vacuum system an electrode assembly to control the operation of the vacuum pumps.

Filter Underdrain System

The filter underdrain system shall be as specified elsewhere. Pressure access hatches to the false filter bottom, if required, shall be furnished by others.

Wash Troughs

Filter wash troughs shall be installed in each filter cell so that they are level throughout and rigidly supported for both internal and external loads. They shall be supplied as indicated on the drawings.

Operating Console

There shall also be furnished and installed on the operating floor, as shown on the plans, an Infilco Operating Console which shall provide remote manual/semi-automatic/automatic control. There shall also be installed in the Console, similar remote controls for operating the surface wash system and valves in each filter cell, if necessary. The details and arrangements of the Console shall be approved by the engineer before fabrication.

Surface Wash (optional)

Surface wash units shall be installed as shown on the drawings and in accordance with the recommendations of the manufacturer of the equipment. There shall be installed in the water supply line to each surface wash unit, a . . . . . . type valve which will be remotely operated by . . . . . . from control valves installed in . . . . . . pipe to washers and their support will be by others.

Filter Isolating Valves (optional)

Filter isolating valves (sluice gates) shall be provided for each control center (filter cell) to isolate the filter units (filter cells) and take them out of service for inspection, maintenance and repair.

INSTALLATION	LOCATION	TOTAL CAPACITY
Nekoosa-Edwards Paper Company	Arkansas Ashdown	17,000
Kimberly Clark Corporation	Connecticut New Milford	4,200
County Water	Florida Ft. Meyers (Lee County)	3,500
Chrysler Corporation	Illinois Belvidere	1,500
Commonwealth Edison Company	Kincaid	1,200
City Water	Indiana South Bend	8,400
Crown Zellerbach Corporation	Louisiana St. Francisville	14,000
City Water	Houma	5,600
City Water	Minnesota Mankato	6,300
City Water	Norwood	400
City Water	White Bear Lake	5,000
City Water	North Dakota Ashley	400
City Water	Dickinson	3,000
City Water	Drayton	500
City Water	Grand Forks	4,200
City Water	South Dakota Springfield	400
E. I. DuPont De Nemours and Company	Texas Beaumont	3,100
Bergstrom Paper Company	Wisconsin Neenah	7,000
City Water	Canada Terrebonne, P.Q.	2,800
City Water	Japan Koriyama	4,400
City Water	Iruma City	4,200
Cable Company, Ltd.	Hitachi	3,500
Fertiberia, S.A.	Spain Huelva	900
Ensidesa Steel, S.A.	Aviles	5,000
City Water	Hong Kong Taipo	5,000
City Water	Viet Nam Bien Hoa	3,000
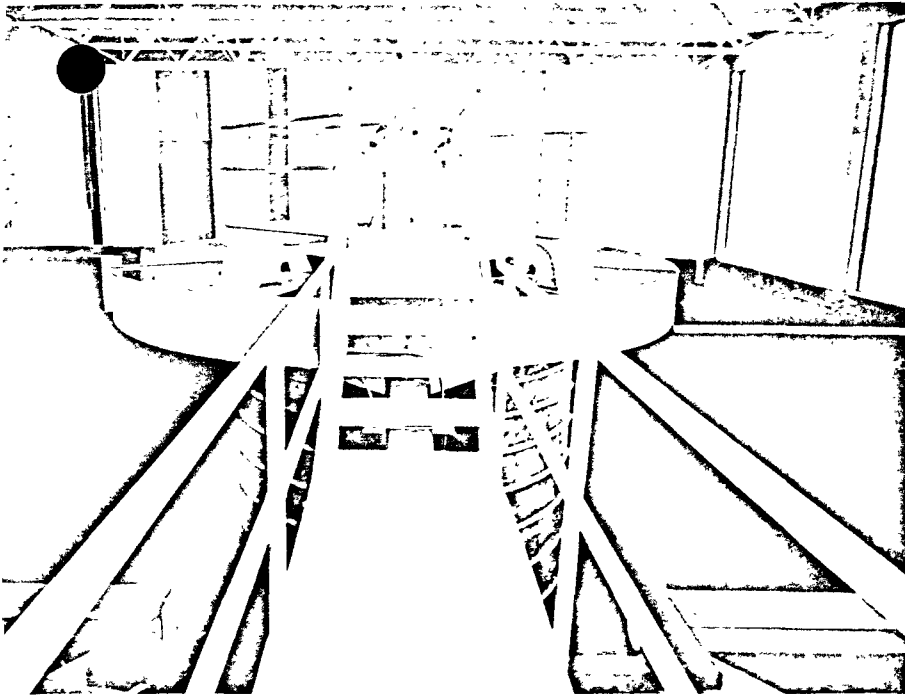
City Water Plant
South Bend, Indiana
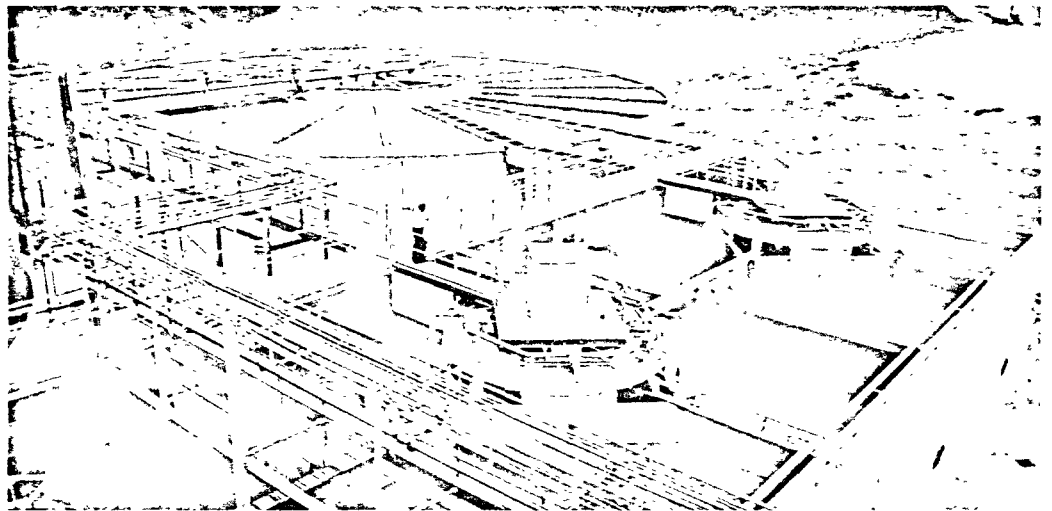
Crown Zellerbach Plant
St. Francisville, Louisiana
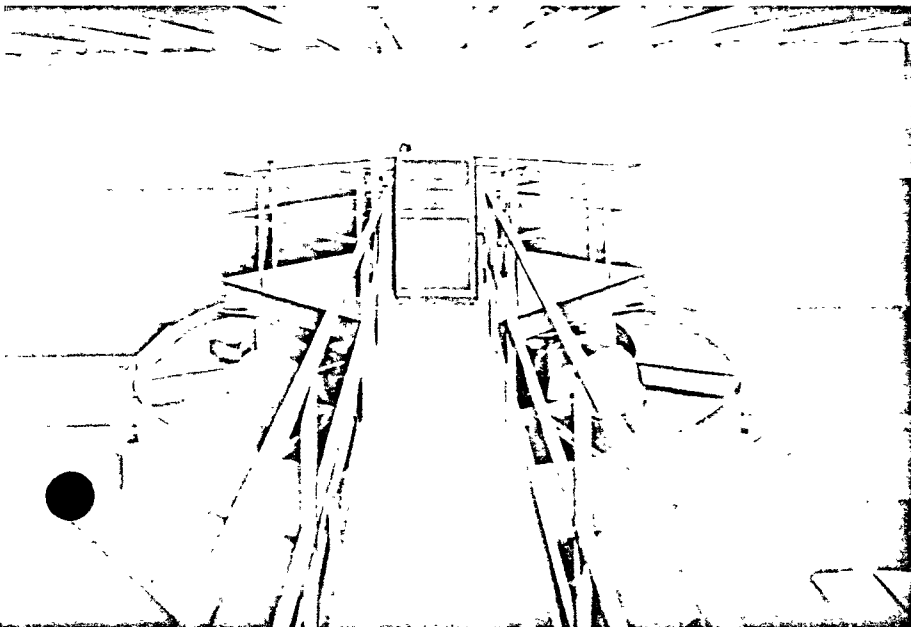
City Water Plant
White Bear Lake,
Minnesota

FULLER
ENGINEERED PRODUCTS FOR INDUSTRY...

AIR HANDLING EQUIPMENT . . . Fuller rotary compressors, Sutorbilt blowers and Lehigh² fans

PNEUMATIC MATERIALS CONVEYING SYSTEMS . . . Airveyor², Airslide² and Fuller-Kinyon conveying systems

POLLUTION CONTROL EQUIPMENT . . . Dracco² dust collectors, Infilco water and sewage treatment systems

HEAVY PROCESSING MACHINERY AND SYSTEMS . . . Traylor crushers, kilns and mills, Fuller grate coolers, fluid bed reactors and Fuller specialized equipment for the mineral industries

Offices in principal cities of the United States and Canada

INFILCO
FULLER COMPANY

P O. BOX 5033, TUCSON, ARIZONA 85703

GATX
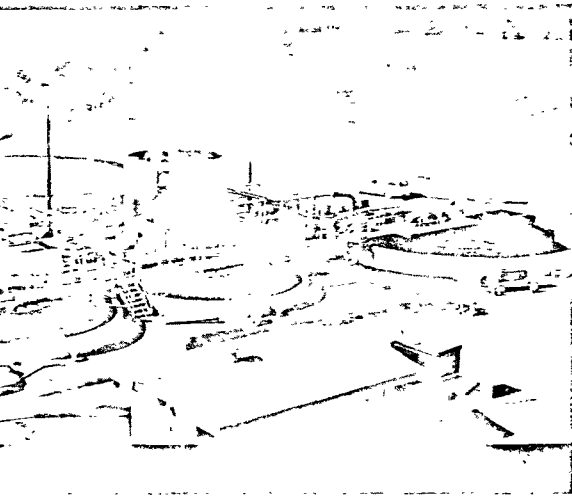
INFILCO

AERO-
ACCELATOR®

activated
sludge
units

What It Is

The "AERO-ACCELATOR" activated sludge plant is proven equipment based upon established principles of hydraulics and bio-chemistry. It is designed for the treatment of both domestic sewage and industrial wastes. The unit provides a compact, highly efficient plant, with conventional design factors established by the Consulting Engineer.

What It Does

The "AERO-ACCELATOR" plant is designed for activated sludge treatment, and incorporates the established advantages of complete mixing. In addition, the "AERO-ACCELATOR" design inherently has plus factors not obtainable in other designs. Consistently, it provides 90% or better BOD removal with rapid and continuous biological oxidation. There is efficient solids separation under positive aerobic conditions.

Why It Should Be Selected

The "AERO-ACCELATOR" plant is of proven design using established loading factors and a highly efficient mechanical air disperser. As a result, it will handle biological loadings in less space than conventional plants. There is consistency of performance and economy in first cost and operating cost.

2

EXPERIENCE

The "AERO-ACCELATOR" activated sludge unit is a proven design. There are more than two hundred and seventy-five successful installations treating domestic sewage and industrial wastes. Consulting Engineers have determined by their experiences with the design and operation of "AERO-ACCELATOR" units, that the consistency of good treatment and its compact economy provide a unit to be specified on many projects.

Experience in the treatment of domestic sewage, phenolic wastes, packing house wastes, paper mill wastes, and others, has been obtained. The largest completely mixed activated sludge plants in the world were both supplied by INFILCO.

The *circular* 175' diameter "AERO-ACCELATOR" unit at Whippany Paper Board Company, Whippany, New Jersey, with a previously supplied 110' diameter unit, handles a total flow of 11 MGD with a BOD of over 400 mg/l from three company mills.

Eight *rectangular* "AERO-ACCELATOR" units at the Shibaura plant in Tokyo are currently treating more than 17 MGD of domestic sewage.

INDEX

Until recently, the design of the activated sludge process has included a great deal of empiricism. Fundamental study of the process has brought about its more efficient and economical use. INFILCO, through its research and development, has pioneered the practical application of the completely-mixed concept and has confirmed biological design factors established by many investigators.*

Activated sludge treatment is a biological process in which bacteria and other types of minute living organisms do the work. If these micro-organisms are to function at maximum efficiency, they should be in a state of constant and uniform growth. The essential conditions of such growth are:

- A continuous supply of waste having a uniform and nutritionally adequate organic content.

- Complete mixing of raw waste and microbial population.

- An uninterrupted supply of dissolved oxygen.

- Efficient separation of treated waste and biological floc so as to retain the latter and discharge the former in a quantity equivalent to the raw waste feed volume.

The traditional system is a cyclic one wherein waste is agitated and aerated in the presence of biologically active floc in an aeration basin. The effluent is subsequently separated by sedimentation, and the floc, or activated sludge, is returned to the process or wasted as required. However, various investigators* have established that there are disadvantages in employing the traditional process with this cyclic system wherein the organisms are over-fed at one stage and then are allowed to become starved and depleted of oxygen before they are reused in the process.

These inherent weaknesses in the traditional activated sludge process have turned the attention of engineers to the development of a homogeneous process.

The completely-mixed system, embodied in the "AERO-ACCELATOR" design provides ideal conditions since the organisms are kept in a constant and uniform state of growth. Incoming wastes are completely mixed with the entire content of the aeration tank. The aeration tank acts as an equalizer of the biological load and the activated sludge is not subjected to shock loadings. Some of the organism which are dying are continually releasing nutrient materials, and in such a homogeneously mixed system, this release and demand for nutrients occur at the same point.

* Note References.

CONTROLLED CONDITIONS

It can be generalized that the efficiency of biochemical treatment, as measured by BOD removal, is related directly to the weight of biologically active solids in the system and inversely to the applied BOD. It follows that neither the waste concentration nor the mixed-liquor aeration period are of fundamental significance. When successful operation at high MLSS (mixed-liquor suspended solids) concentrations is possible, considerably smaller aeration basins can be used than have been utilized in the past. Economic and technical limitations in this regard include the oxygenation capacity of the aeration system, the influence of MLSS concentration on the effectiveness of liquid-solids separation, and the return-sludge capacity required for operation at high MLSS levels.

The oxygen absorption efficiency of conventional diffused-air systems ranges up to 10% and most of the air supplied is required to keep the sludge in suspension. Experience has demonstrated that the VORTI-MIX® turbine aerator (or VORTAIR® surface entrainment aerator) component of the "AERO-ACCELATOR" unit will properly mix and oxygenate higher MLSS than can be effectively separated in conventional practice.

The circulation between the aeration and clarification zones can be controlled. In most applications, it is four to five times the throughput

This is several times the return-sludge capacity provided in the traditional system, and eliminates this particular restriction on MLSS concentrations.

Liquid-solids separation in the traditional activated sludge process is accomplished by transferring mixed-liquor solids from the aeration basin to a gravity-type settling and clarification unit. Due to the time required for settling and thickening, the biological solids are kept away from any dissolved oxygen for as much as an hour or more. This is undesirable since aerobic organisms cannot maintain their vitality for any length of time in the absence of dissolved oxygen.

In the "AERO-ACCELATOR" design, there is no possibility of sludge septicity caused by settling outside the zone of active aeration In the "AERO-ACCELATOR" unit, biochemical oxidation and clarification take place in a single multi-purpose structure. The activated sludge is constantly being recirculated, under aerobic conditions, from the solids separation zone into the aeration zone.

Clearly, the "AERO-ACCELATOR" plant provides a biological environment which meets the requirements outlined for an efficent treatment system.
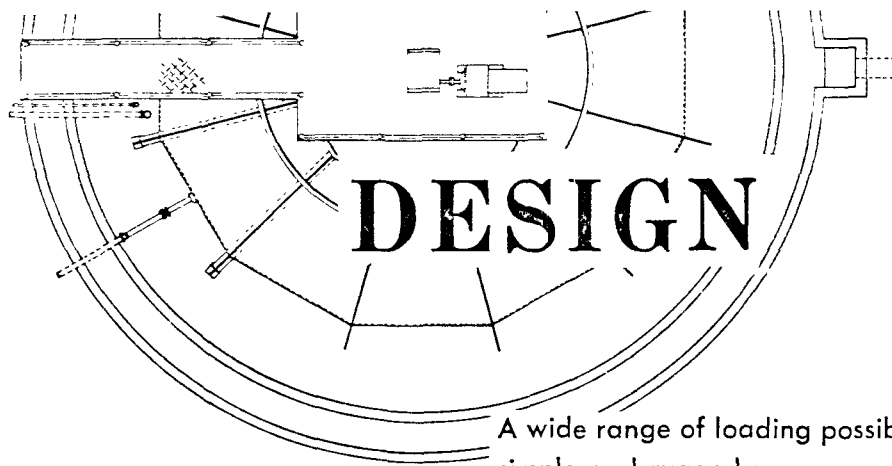
DESIGN

A wide range of loading possibilities is available for a specific design of
simple and rugged process as applied in the "AERO-ACCELATOR" u
The engineer may design an extended aeration system from which littl
no sludge is wasted; he may elect to design for operation at a higher BC
MLSS ratio with consequent excess sludge production. The choice is influe
ed by local requirements and conditions.

As stated by one investigator*, "The desired effluent quality determines the size
the complete mixing system. It is possible to produce an effluent of any desi
organic level from wastes of any organic strength."

BASIC FACTORS

The basic design factors that establish the size and volume of an "AEF
ACCELATOR" unit are the aeration volume, total mixed-liquor volume necess
for handling the biological load, and the clarification area necessary for so
separation.

LOADINGS

Data obtained under controlled conditions indicate that for the completely-mi
system as applied in the "AERO-ACCELATOR" unit, BOD/MLSS loading re
considerably greater than in the traditional activated sludge process are enti
practicable for the same degree of treatment. However, for domestic sew
treatment at an average loading of 0.5 pound of BOD per day per pound of N
is recommended. Basic unit dimensions on pages 18 to 21 provide for conserva
and conventional loadings incorporating an appreciable safety factor for the st
normal conditions. Higher concentrations of MLSS can usually be maintained,
sizing of the "AERO-ACCELATOR" unit incorporates adequate high-efficie
oxygenation to sustain any economical loading condition. For stronger was
loading ratios in the range 0.5 to 1.0 and higher are often practicable.

CLARIFICATION

The clarification area is established on the basis of the settling characteristics of
mixed liquor. For ordinary activated sludge developed from domestic sewag
maximum overflow rate of 1500 g.p.d. per sq. ft. is used.

AIR REQUIREMENTS

Air requirements are determined from the BOD loading, the oxygen required
pound of BOD removed, and the oxygen transfer efficiency of the aeration syst
The oxygenation device in the "AERO-ACCELATOR" unit is a non-clogging
bine aerator which either disperses compressed air or entrains atmospheric
Oxygen absorption efficiencies of 15 to 25% are economical. The "VORTI-M
aerator provides about 2.5 pounds of oxygen per hour per horsepower appl
including power for air compression. The "VORTAIR" aerator will provide u
about 5.0 pounds of oxygen per hour per horsepower applied to the rotor.

6

* Note Reference 8

The "AERO-ACCELATOR" unit offers you —

Savings up to 50% in space requirements.

Mixing, oxygenation and clarification in a single unit.

Round, square or rectangular designs.

"Extended Aeration" if required.

Most efficient biological oxidation in the most compact plant.

STANDARD FLOW SHEETS
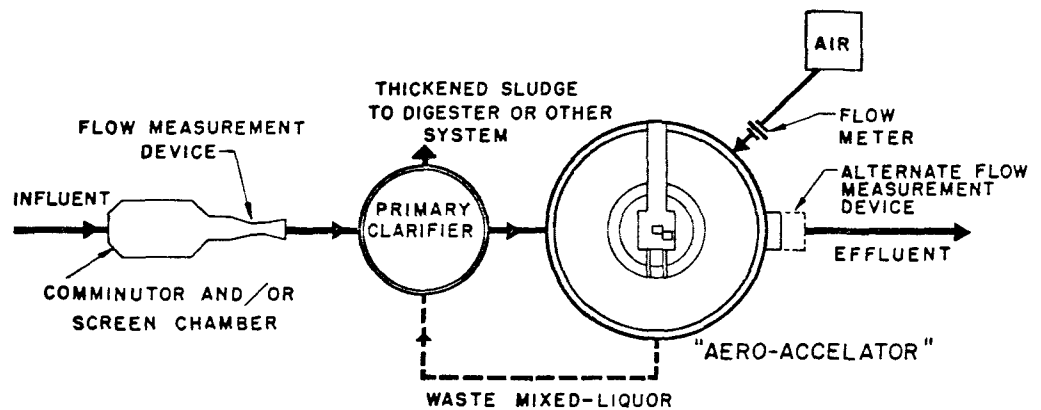
1. FLOW DIAGRAM FOR NORMAL DOMESTIC SEWAGE.
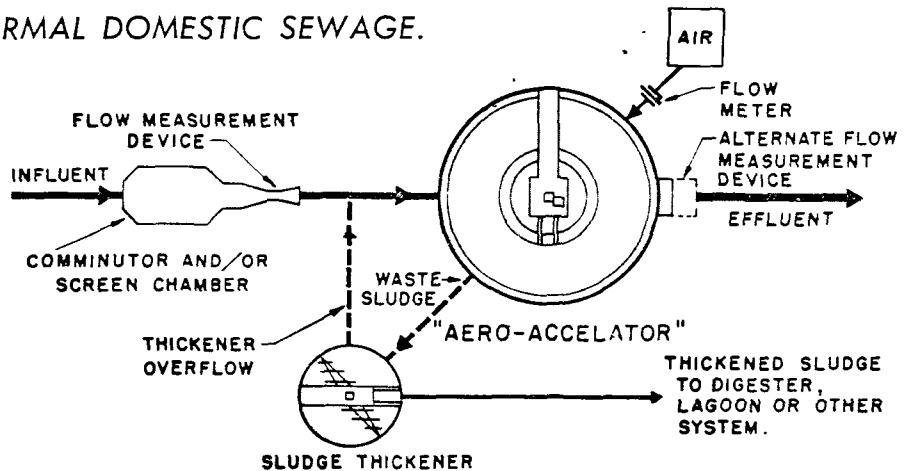
2. FLOW DIAGRAM FOR WASTE WATERS NOT REQUIRING PRESETTLING.
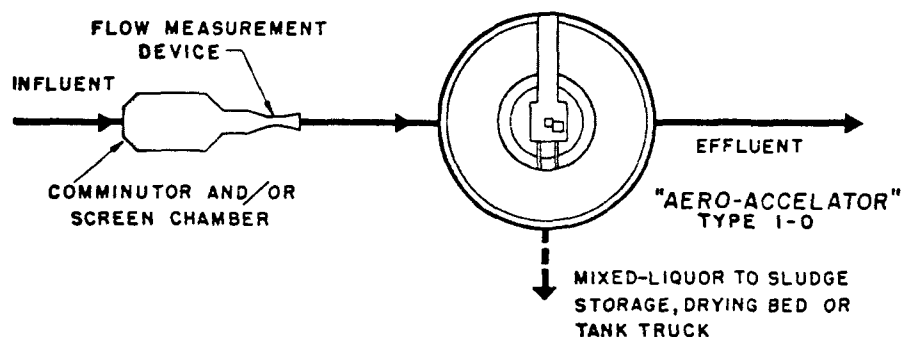
3. FLOW DIAGRAM FOR APPLICATION OF TYPE I-O, FOR EXTENDED AERATION.

The"AERO-ACCELATOR" Treating Plant with its controlled mixed liquor circulation provides the smallest completely mixed activated sludge plant available for the effective treatment of municipal and industrial wastes.

The"AERO-ACCELATOR" unit is designed for the treatment of domestic sewage.

The "AERO-ACCELATOR" plant is also designed for the treatment of paper mill, refinery, and other complex chemical wastes.

The following diagrams and explanations will describe the flow and functions of the "AERO ACCELATOR" plant design:

Waste enters the "AERO-ACCELATOR" —
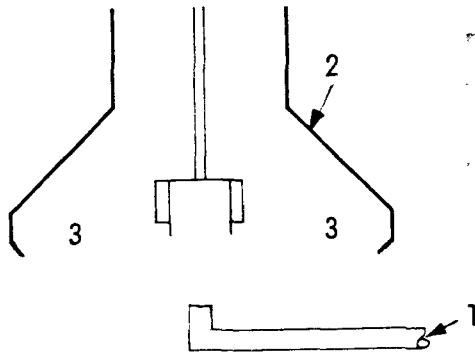
Figure 4

The waste flows through the intake pipe (at the bottom center below the hood (2) in the aeration and mixing zone (3).
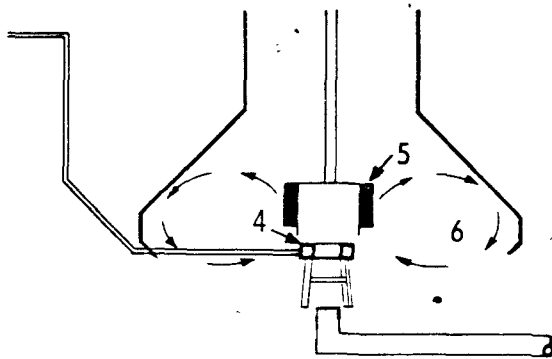
Figure 5

At this point, air is introduced through a spar ring (4) and is dispersed by the"VORTI-MI Aerator (5).

The"VORTI-MIX" Aerator also acts as a rad type pump which completes mixing of the and waste and circulates the mixtures with the aeration zone (6).
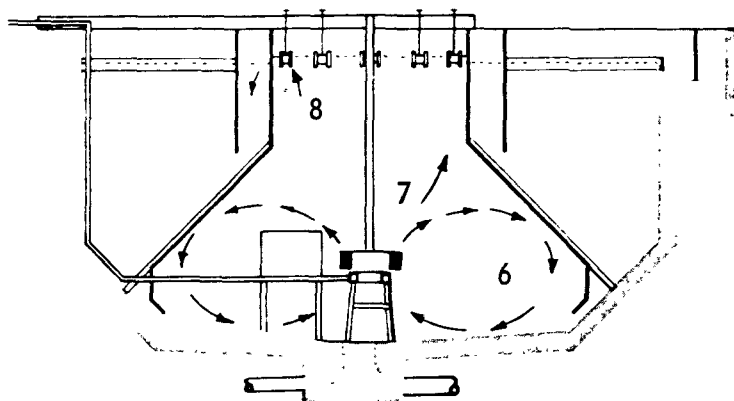
Figure 6

The large bubbles of incoming air are brok into very fine bubbles by the shearing-acti of the rotor. These fine bubbles promote ra oxygenation and provide a lifting action which conveys the mixed liquor up the in draft tube where the flow discharges throu variable opening gates (8).

8

The mixed liquor then moves down the annular space between draft tubes (9) to the clarification zone (10). A throughput volume of clarified effluent is displaced from the mixed liquor by the newly entering waste. The effluent rises and flows over the discharge weir (11) at the surface.
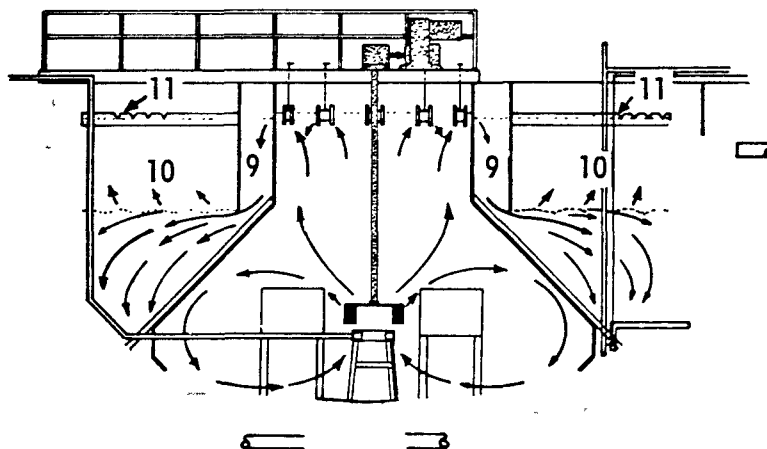
Figure 7

The activated solids separated from the influent are carried back into the mixing zone beneath the hood by the recirculating volume of mixed liquor (12). The recirculating flow is normally four or five times the throughput and is controlled by varying the area of the gate openings (8) Solids concentration is controlled by a timer-activated blowdown (13) or an air-lift (14) to an external thickener.
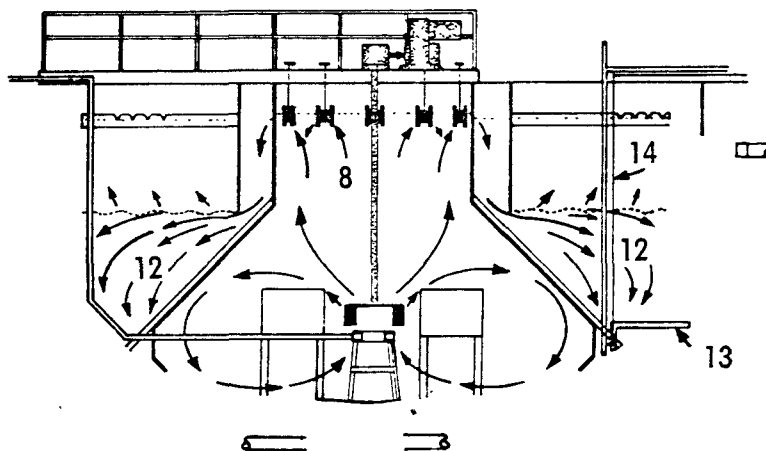
Figure 8

Type I-O

When designing for high-strength industrial wastes, where the mixed-liquor volume required is relatively large in comparison with the clarification area needed, the "AERO-ACCELATOR" unit TYPE I-O is used. In this unit, described in the section — "Modifications" page 14, the positions of the aeration and clarification zones are reversed.

The "AERO-ACCELATOR" activated sludge plant TYPE I-O was designed for the treatment of wastes of high BOD content. Many industrial wastes, which often have BOD contents of more than 600 mg/l require more aeration of a greater volume of mixed liquor than that required for domestic wastes The design of the TYPE I-O unit provides this additional volume within the aeration zone by placing the larger aeration zone at the periphery of the basin and the smaller clarification zone within it. The operation of the TYPE I-O is similar to that of the Standard unit, therefore, the same identifying numerals apply.
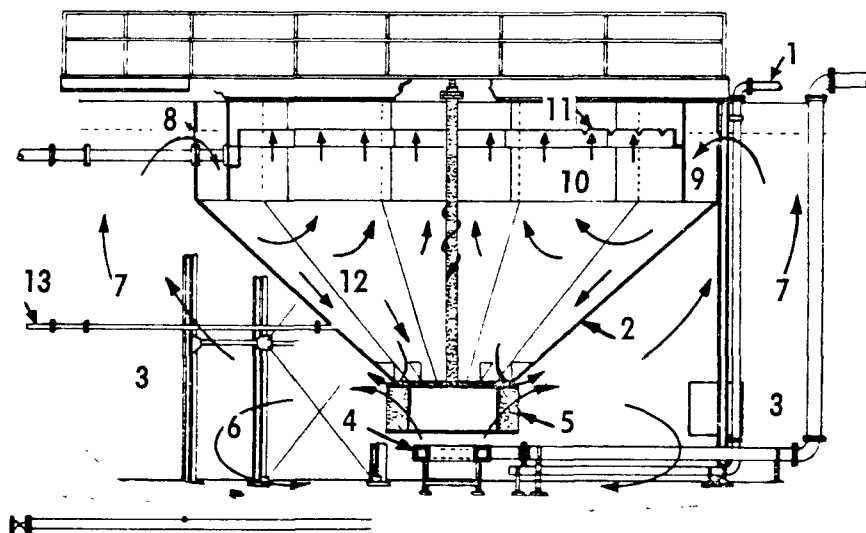
Figure 9

Analytical data from a variety of full-scale installations and pilot plant studies are available. The treatment of several different types of wastes is illustrated.

DOMESTIC SEWAGE — AVERAGE HOUSING

Run No.	BOD			SUSPENDED SOLIDS		
	Influent (mg/l)	Effluent (mg/l)	Removal (%)	Influent (mg/l)	Effluent (mg/l)	Removal (%)
1	151	15	90.0	147	16	88 9
2	127	6	94.5	147	22	86 5
3	168	7	95 6	141	15	89 7
4	157	7	95.4	125	13	89 6
5	134	7	94.7	120	16	87.3
6	149	6	95.7	166	16	90 3
7	141	6	95 6	81	8	90 1
8	111	5	95 8	74	5	93 2
9	125	10	90 8	89	13	85.0
10	171	16	91.0	176	21	88 0
11	219	14	93 7	181	13	92 8
12	158	9	93 6	121	8	87 0
13	158	10	93.6	126	9	91 0
14	126	10	91.8	92	11	85.0
15	156	8	94.8	104	7	93.3
16	147	3	97.7	160	10	93.2
17	125	5	96.3	119	10	91.7
18	87	2	96 7	79	5	93 7
19	94	3	97 2	85	4	96.0
20	96	3	97 0	87	6	92 8
21	99	5	95.0	80	6	92.5

The flow rate to the plant varied from 20 g p m to 110 g.p m during the test period.

Figure 10

DOMESTIC SEWAGE — LARGE MOTEL

Period	BOD			SUSPENDED SOLIDS		
	Influent (mg/l)	Effluent (mg/l)	Removal (%)	Influent (mg l)	Effluent (mg/l)	Removal (%)
Jan. 1955	187	10	94	383	38	90
Feb	164	12	93	408	32	92
March	137	7	95	383	28	93
April	211	25	88	328	38	88
May	171	9	94	222	9	96
June	173	8	95	245	32	89
*July	174	49	72	228	101	56
*Aug.	145	30	79	221	65	70
Sept	—	—	—	—	—	—
Oct.	127	10	92	147	24	84
Nov.	264	7	97	218	7	97
Dec.	169	11	93	260	20	92
Jan. 1956	216	18	91	—	—	—
Feb.	200	17	91	228	37	84
March	222	13	94	227	56	75
April	190	16	91	314	65	80
May	164	15	91	149	· 8	94
June	171	13	92	165	32	80
July	152	4	97	205	41	80
Aug.	136	8	94	185	9	95
Sept	120	10	91	241	25	89
Oct.	149	21	86	188	38	80
Nov.	166	11	93	162	28	83
Dec.	175	12	93	232	30	87

Sewage is pumped from a surge tank at 30 g p m Plant operation is intermittent.
*"VORTI-MIX" Aerator out of service

Figure 11

PHENOLIC WASTE WATER — OIL REFINERY

Period	PHENOL			SULFIDES		
	Influent (mg/l)	Effluent (mg/l)	Removal (%)	Influent (mg /l)	Effluent (mg /l)	Removal (%)
Jan. Feb 1956	180	0.4	99.8	40	0	100%

Date	BOD			SUSPENDED SOLIDS		
	Influent (mg/l)	Effluent (mg/l)	Reduction (%)	Influent (mg/l)	Effluent (mg/l)	Reduction (%)
9-30-54	264	9 2	96 5	117	7.7	43.4
10-25-54	321	24 0	92 5	301	16 6	94 5
10-30-54	325	21.0	93 6	274	15 7	94.3
11-16-54	271	20 5	92.4	268	21.8	91.9
3- 4-55	169	14.0	91 7	125	11 0	91.2
7-30-59	1830	69 0	96 2	4145	19.0	99 5
*10- 1-59	593	44.0	92 6	1016	33 0	96.7
**10- 1-59	593	36.0	93 9	1016	21.0	97.9

*Raw Sewage and Effluent of "AERO-ACCELATOR" No. 1
**Raw Sewage and Effluent of "AERO-ACCELATOR" No. 2

Figure 13

CONSISTENTLY HIGH REMOVALS
OVER WIDE RANGE OF LOADINGS
WITH LOW EFFLUENT BOD

by

the

"AERO-ACCELATOR"
unit
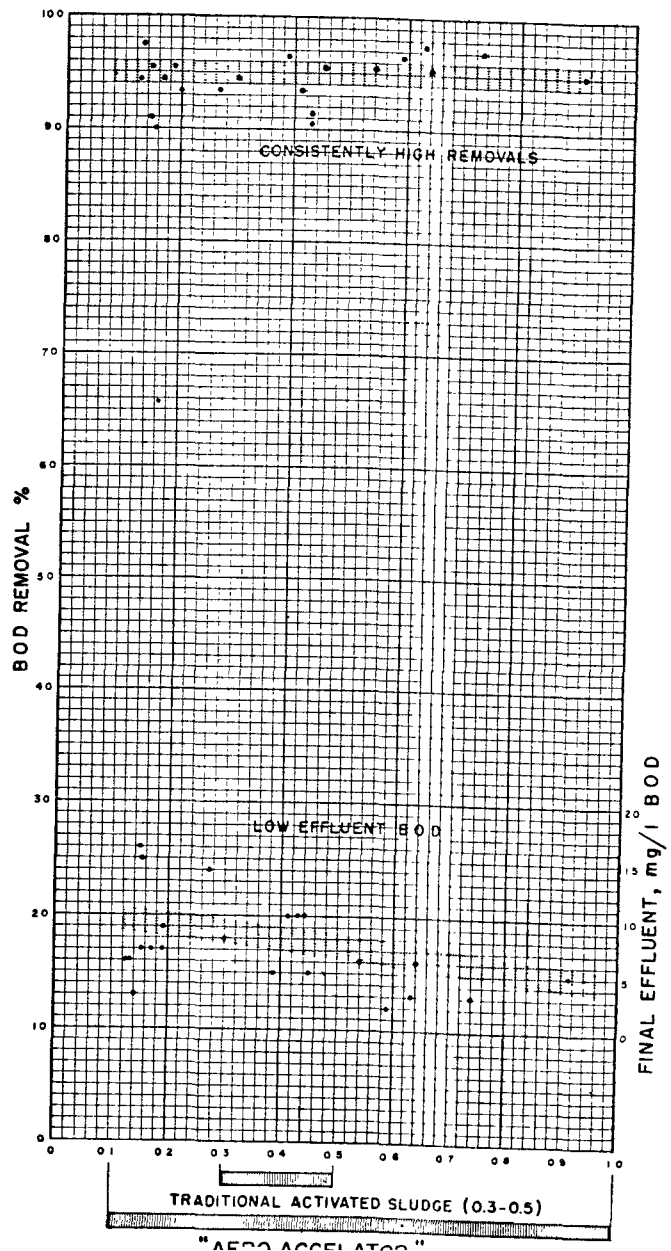
● ● ● The proper application of an "AERO-ACCELATOR" unit will depe
upon the strength and variability of the waste to be treated, and on t
results desired. Required design information includes the average a
peak flow rates, waste characteristics (BOD, suspended solids, pH, te
perature, and other criteria), and the desired effluent quality. The
data determine the necessary aeration and mixed liquor volumes wh
are based upon experience with numerous pilot and full-scale plan

● ● ● Provision of proper aeration volume and limitation of clarificat
area loading define the application of the "AERO-ACCELATO
unit. Detention period, in itself, is of no significance. Different si
of internal mechanisms can be installed in the same size bas
in order to accommodate both the aeration and clarification requ
ment of a specific treatment problem.

● ● ● Requirements likewise dictate the oxygenation capacity of
aeration unit, whether the "VORTI-MIX" or the "VORTAIR" ae
tor is selected. The design and flexibility in operation of each ae
tor maintain mixed-liquor dissolved oxygen at minimum t
power consumption. Because adequate mixing is provided by
rotor, no dependence is placed upon air to accomplish mixing.

MODIFICATIONS

The "AERO-ACCELATOR" unit is offered with certain basic modifications
adapt to a wide variety of conditions and wastes.

The basic unit has been described under "Operation" and can be applied for norr
activated sludge treatment of domestic sewage or wastes with a BOD of less tl
about 600 mg/l.

For stronger wastes, or for treatment of domestic sewage by "extended aeratic
the standard unit is modified so the geometrical proportions will provide the n
economical unit consistent with the waste and the process. In the modificat
TYPE I-O, the positions of aeration and clarification zones, are reversed.
change provides the required large aeration volume and relatively small clarifica
area. By this means, loadings several times those permissible for the stand
"AERO-ACCELATOR" unit can be used without sacrificing any of the dem
strated advantages of effective activated sludge treatment. Strong wastes can
treated effectively and economically. The overall area required is considerably
than that of other comparable treatment systems.

Figure 9, Page 9, in the section under "Operation", illustrates the "AER(
ACCELATOR" unit, TYPE I-O.

Type S

For sewage treatment, the "AERO-ACCELATOR" unit is preceded by a primary clarifier with scum and sludge removal mechanisms. When the clarifier is designed according to customary standards of engineers and regulatory agencies, advantage is taken of economical BOD removal to decrease the loading and size of the "AERO ACCELATOR" unit. Waste mixed-liquor solids can be returned to the primary clarifier for thickening prior to sludge disposal.

Since the settleable or floatable solids content of many industrial wastes is either very low or non-existent, pre-treatment is not always necessary. Such wastes can be discharged directly to the "AERO-ACCELATOR" plant. However, if previous experience does not provide the necessary design information to establish all load factors for efficient treatment, they should be established by pilot plant study.

Mechanisms are available for installation in either circular or square basins.

Type I-O
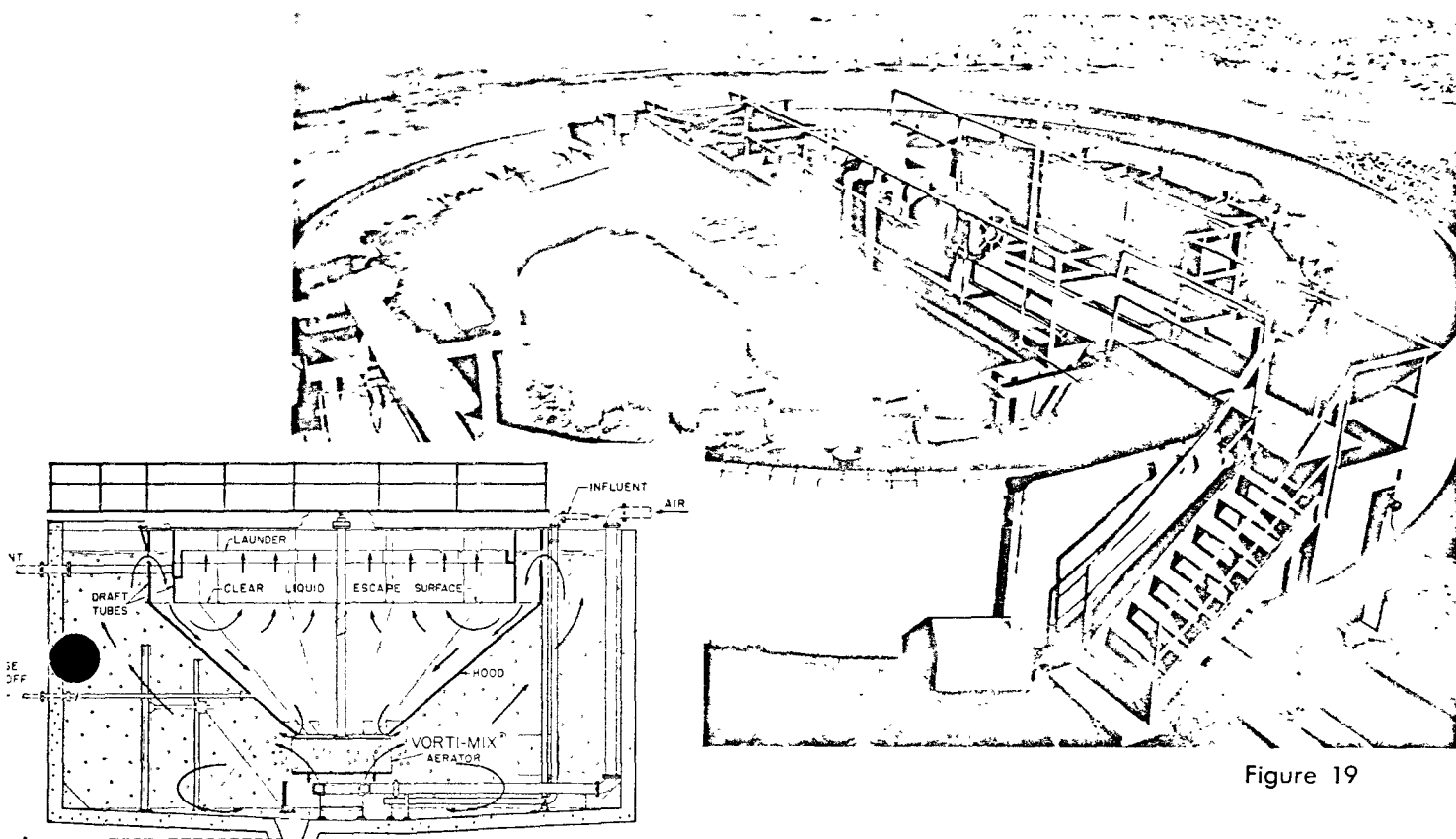
Figure 19
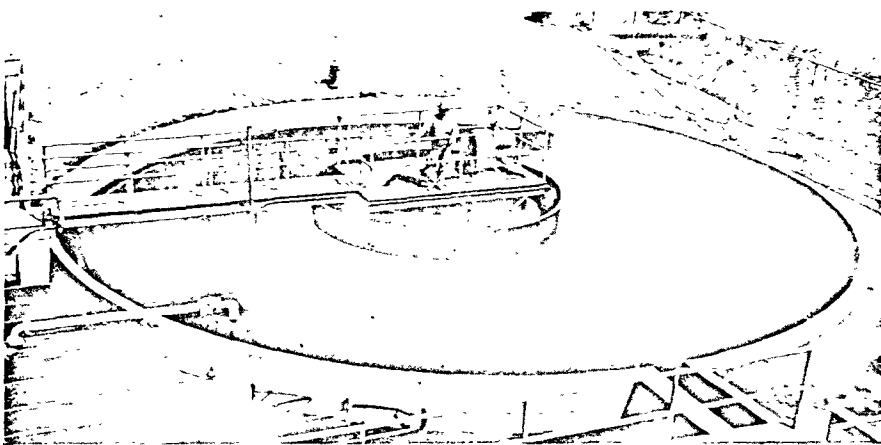
City of Elliot Lake
Elliot Lake, Ontario
Canada
45-Foot Diameter

Figure 20

Whippany Paper Board Com
Whippany, New Jersey
110-Foot Diameter

Figure 21
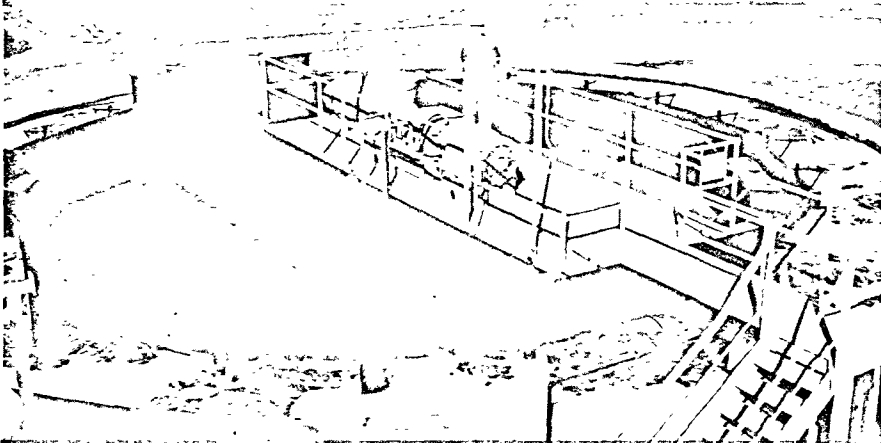
Mobil Oil Company
Ferndale, Washington
(2) 28-Foot, Standard
(1) 50-Foot, Type I-O

Figure 22
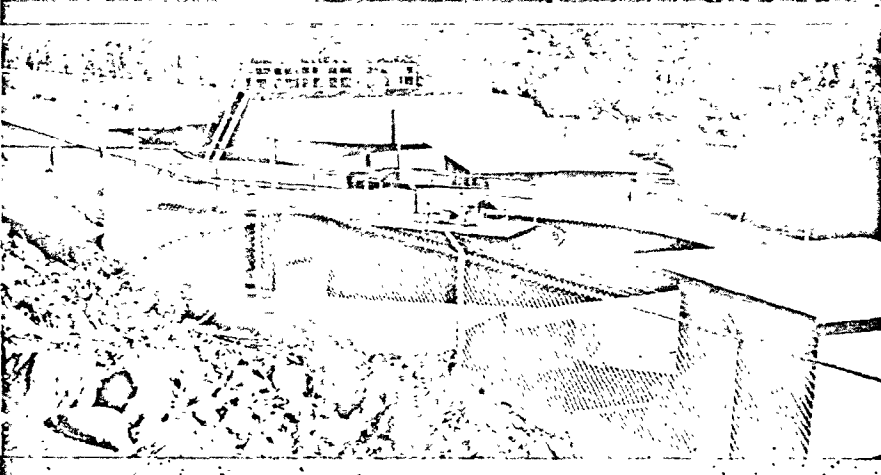
Kaanapali Development
Maui, Hawaii
47-Foot Diameter

Figure 23

16

Domestic and Industrial Plants

Shibaura Sewage
Tokyo, Japan
(8) 52-Feet Wide By
     131-Feet Long
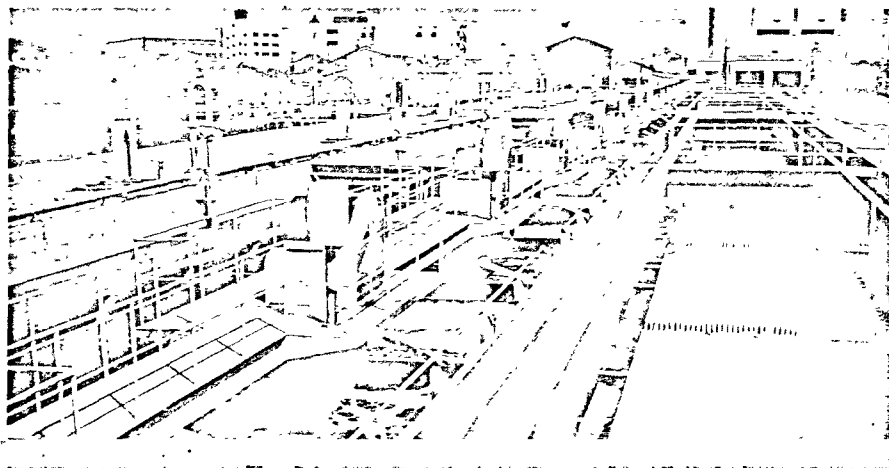
Figure 24

Whippany Paper Board Company
Whippany, New Jersey
75-Foot Diameter
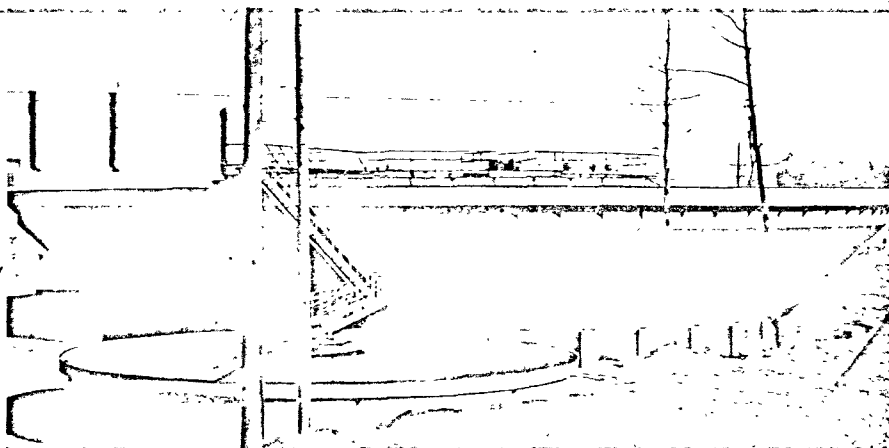
Figure 25

City Sewage
Schwaebisch, Gmünd
Germany
(2) 53-Foot Diameter
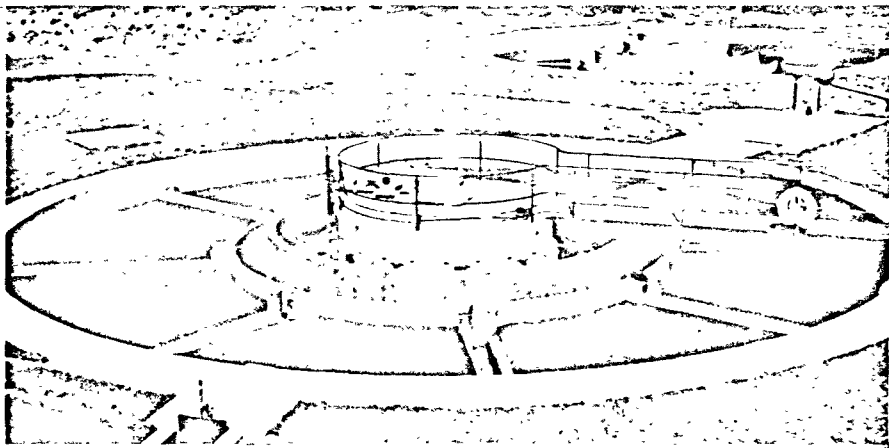
Figure 26

City of Litchfield
Litchfield, Minnesota
(2) 47-Foot Diameter
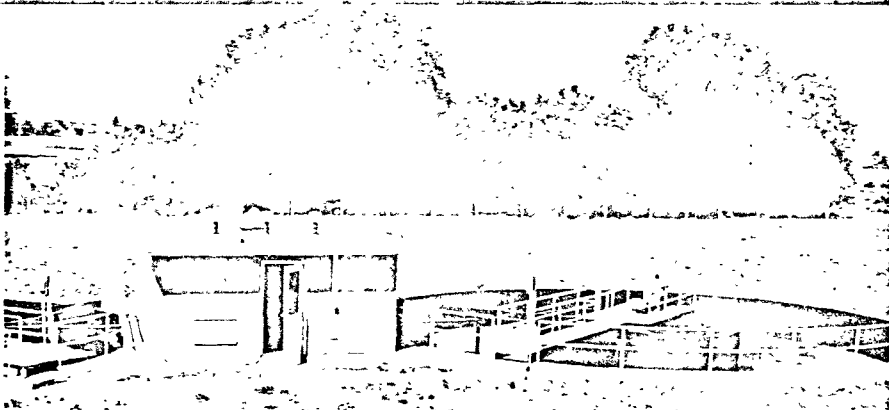
Figure 27
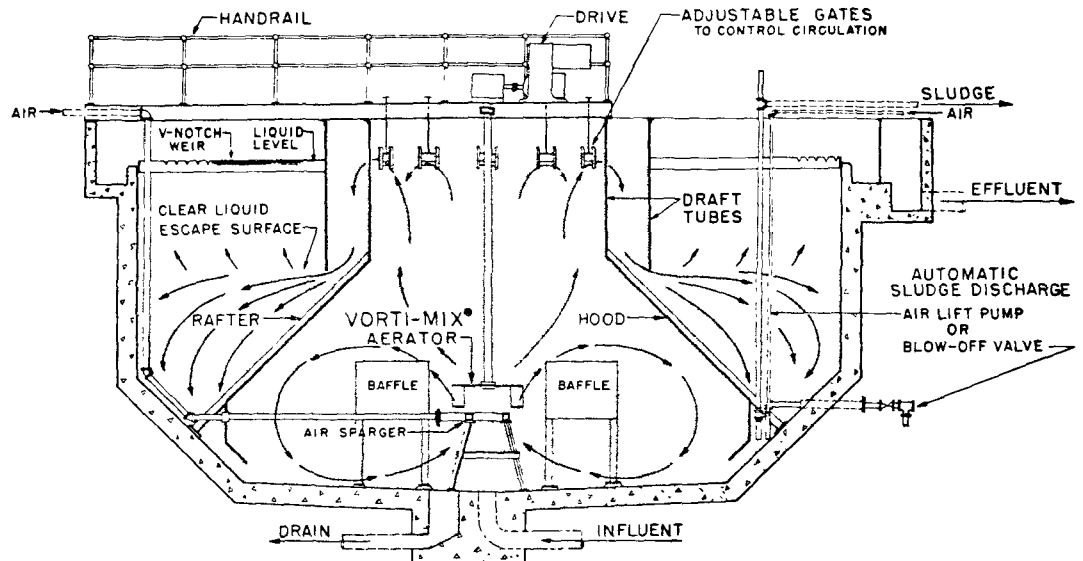
Figure 28

APPROXIMATE DIMENSIONS FOR STANDARD

Based on these assumed conditions· 200 mg/l BOD in Raw Sewage • 30% BOD removal in Primary • 12% Oxygen Absorption Efficiency • Maximum Flow = 200% of Avg. •Temp of waste = 29°C • Mixed Liquor DO = 2 Mg/l • O Saturation Value = 95%, that of water • O. Transfer Factor, α, = 0.95.

Average Flow g.p.d.	Basin Dimensions				Clarification Area sq. ft.	Total Volume gal.
	Diameter	S.L.D.	Freeboard	Diam. at Toe of Fill		
100,000	15'-0"	11'-6"	1'-6"	10'-0"	144	14240
150,000	19'-0"	11'-6"	1'-6"	10'-0"	250	20600
200,000	21'-0"	12'-0"	1'-6"	12'-0"	302	26820
300,000	25'-0"	13'-0"	2'-0"	14'-6"	420	40950
400,000	29'-0"	13'-3"	2'-0"	16'-6"	566	54050
500,000	33'-0"	13'-6"	2'-0"	18'-6"	723	69000
600,000	36'-0"	14'-0"	2'-0"	20'-6"	864	85900
700,000	38'-0"	14'-0"	2'-0"	20'-6"	980	90000
800,000	42'-0"	15'-0"	2'-0"	24'-0"	1184	121200
900,000	44'-0"	15'-0"	2'-0"	24'-0"	1320	126800
1,000,000	47'-0"	16'-0"	2'-0"	26'-6"	1467	158000
1,250,000	52'-0"	16'-6"	2'-0"	30'-0"	1777	199000
1,500,000	57'-0"	17'-0"	2'-0"	33'-0"	2118	242000
1,750,000	62'-0"	18'-0"	2'-0"	36'-6"	2528	304000
2,000,000	67'-0"	18'-6"	2'-0"	42'-0"	2818	381000
2,250,000	72'-0"	19'-0"	2'-0"	45'-0"	3242	444000
2,500,000	77'-0"	19'-3"	2'-0"	51'-6"	3493	534000
2,750,000	80'-0"	19'-3"	2'-0"	51'-6"	3862	555000
3,000,000	84'-0"	20'-0"	2'-0"	55'-0"	4253	645000

18
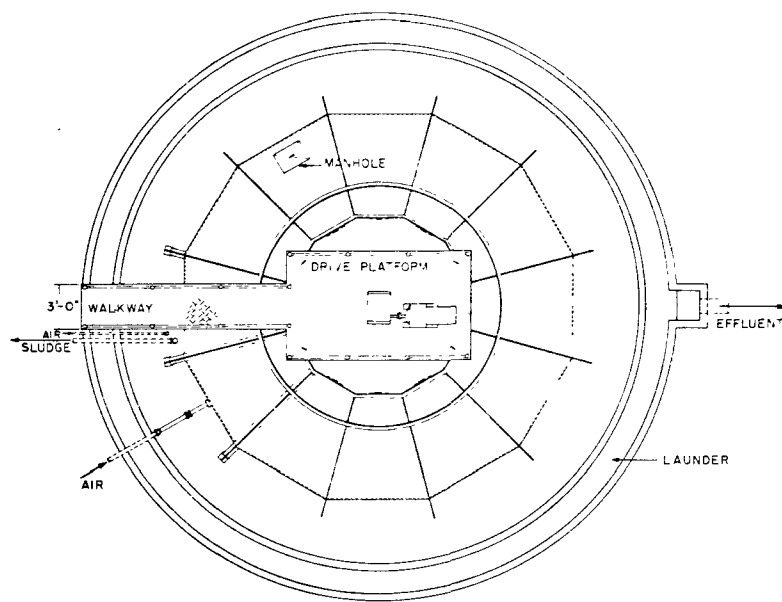
PLAN

Figure 29

\ERO-ACCELATOR" UNIT

Mixed Liquor Volume Gal.	@ 3,000 mg/l MLSS		Aerator Drive HP	Air Required (Not including Standby)			Average Flow g.p.d.
	lbs. MLSS	lb. BOD / day /lb. MLSS		Volume scfm	Pressure psig	Blower HP	
9940	250	0.47	1	32	5.0	2	100,000
13100	330	0.54	1½	48	5.0	3	150,000
17760	440	0.53	2	64	5.5	3	200,000
26790	670	0.52	3	96	5.5	5	300,000
35000	880	0.54	5	128	6.0	7½	400,000
44650	1120	0.52	7½	160	6.0	7½	500,000
56800	1420	0.49	7½	192	6.0	7½	600,000
57000	1420	0.57	7½	220	6.0	10	700,000
82100	2050	0.46	10	256	6.0	10	800,000
82300	2060	0.51	10	286	6.0	10	900,000
103150	2580	0.45	15	320	6.5	15	1,000,000
132000	3310	0.44	15	400	6.5	15	1,250,000
162000	4060	0.43	15	480	7.0	20	1,500,000
210000	5250	0.39	20	560	7.0	30	1,750,000
2~~000	6820	0.34	25	640	7.5	40	2,000,000
3~~000	8060	0.33	25	720	7.5	40	2,250,000
403000	10100	0.29	30	800	7.5	40	2,500,000
410000	10300	0.31	40	880	7.5	40	2,750,000
485000	12140	0.42	40	960	8.0	50	3,000,000

"AERO-ACCELATOR" unit
with "VORTAIR" Aerator

APPROXIMATE DIMENSIONS F

Based on these assumed conditions: 200 mg/l BOD in Raw Sewage • 30% BOD
removal in Primary • Maximum Flow = 200% of Avg. • Temp. of waste = 29°C
• Mixed Liquor DO = 2 Mg/l • O₂ Saturation Value = 95%, that of water • O₂ Transfer Factor, α, = 0.95.

Average Flow g.p.d.	Basin Dimensions			
	Diameter	S.L.D.	Freeboard	Diam. at Toe of Fill
100,000	16'-0"	10'-0"	2'-0"	8'-6"
150,000	19'-0"	10'-0"	2'-0"	10'-0"
200,000	22'-0"	10'-0"	2'-0"	11'-6"
300,000	26'-0"	11'-0"	2'-0"	14'-6"
400,000	30'-0"	11'-0"	2'-0"	16'-0"
500,000	35'-0"	12'-0"	2'-6"	20'-0"
600,000	39'-0"	12'-0"	2'-6"	24'-0"
700,000	45'-0"	13'-0"	3'-0"	32'-0"
800,000	47'-0"	13'-0"	3'-0"	32'-0"
900,000	49'-0"	13'-0"	3'-0"	32'-0"
1,000,000	54'-0"	13'-0"	3'-0"	37'-6"
1,250,000	60'-0"	14'-0"	3'-0"	42'-6"
1,500,000	65'-6"	15'-0"	3'-0"	47'-0"
1,750,000	74'-0"	15'-0"	3'-6"	55'-0"
2,000,000	78'-0"	16'-0"	3'-6"	58'-0"
2,250,000	83'-6"	16'-0"	3'-6"	63'-0"
2,500,000	87'-6"	17'-0"	3'-6"	66'-0"
2,750,000	90'-0"	17'-0"	3'-6"	66'-0"
3,000,000	93'-0"	18'-0"	4'-0"	69'-0"
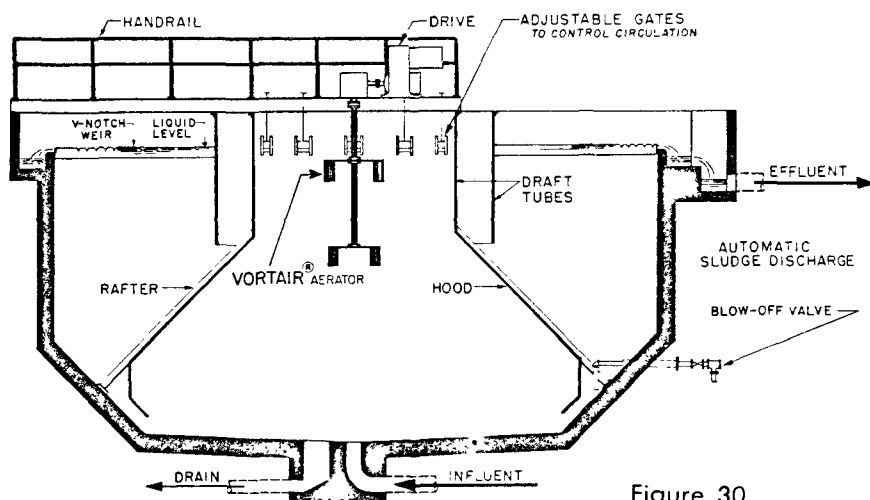
Figure 30

STANDARD "AERO-ACCELATOR" UNIT

Clarification Area sq. ft.	Total Volume gal.	Mixed Liquor Volume Gal.	@ 3,000 mg/l MLSS		Aerator Drive HP	Average Flow g.p.d.
			lbs. MLSS	lb. BOD /day /lb. MLSS		
157	12850	8150	205	0.58	3	100,000
220	17350	10780	270	0.65	5	150,000
294	22400	13650	340	0.69	5	200,000
418	35100	22600	575	0.61	7½	300,000
564	43600	26750	670	0.70	10	400,000
748	66500	41400	1035	0.57	10	500,000
848	85000	56300	1410	0.50	15	600,000
1018	134500	100000	2500	0.33	20	700,000
1163	141000	102000	2550	0.37	20	800,000
1313	146200	102200	2560	0.41	20	900,000
1435	183500	135500	3390	0.34	25	1,000,000
1953	248000	182000	4550	0.32	30	1,250,000
2113	306000	227000	5670	0.31	30	1,500,000
2491	412000	319000	8000	0.26	40	1,750,000
54	487500	380000	9500	0.25	50	2,000,000
3144	560000	444000	11100	0.24	50	2,250,000
3550	656000	524000	13100	0.22	50	2,500,000
3898	670000	525000	13200	0.24	60	2,750,000
4242	773000	614000	15300	0.23	75	3,000,000

● ● ● ● From the 275 successful installations, the following are typ of the acceptance of "AERO-ACCELATOR" units. The wide-spr geographical locations with designs by many consulting engineers, various types of sewage and industrial wastes, and the utilization proper modifications, all attest to acceptance by progressive engine

Location		Size	Waste	Type
Arkansas, Sherwood — City of Sherwood		30'-0" dia. x 16'-6" SLD	Sewage	I-O
California, Edgemont — Edgemont Community Service District		35'-0" dia. x 13'-6" SLD	Sewage	Std.
Florida, Broward County — Cooper Colony Estates		30'-0" dia. x 14'-0" SLD	Sewage	Std.
Hawaii, Island of Maui — Kaanapali Development		47'-0" dia. x 16'-0" SLD	Sewage	Std.
Illinois, Frankfort — City of Frankfort		23'-0" dia. x 13'-3" SLD	Sewage	Std.
Kentucky, Louisville — Whispering Hills Subdivision		30'-0" dia. x 13'-3" SLD	Sewage	Std.
Massachusetts, Sunderland — City of Sunderland		29'-0" dia. x 16'-6" SLD	Sewage	I-O
Michigan, Kalamazoo — Kalamazoo Paper Company		42'-0" dia. x 19'-6" SLD	De-inking Waste	I-O
Minnesota, Litchfield — City of Litchfield	(2)	47'-0" dia. x 16'-0" SLD	Sewage	Std.
Minnesota, Rose Port — Great Northern Oil Company		60'-0" dia. x 18'-3" SLD	Oil Refining Waste	Std.
Nebraska, Chappell — City of Chappell		25'-0" dia. x 13'-6" SLD	Sewage	Std.
New Jersey, Hamilton Township — Yardville — Groveville Plant	(2)	28'-0" dia. x 12'-6" SLD	Sewage	Std.
New Jersey, Whippany — Whippany Paper Company		110'-0" dia. x 22'-6" SLD	Board Mill Waste	Std.
		175'-0" dia. x 29'-0" SLD		Std.
New York, Monticello — Lake Louise Marie	(2)	29'-0" dia. x 13'-3" SLD	Sewage	I-O
North Carolina, Charlotte — Lance Inc.		31'-0" dia. x 16'-6" SLD	Sewage	I-O
North Carolina, Shelby — Pittsburgh Plate Glass Co.	(2)	35'-0" dia. x 20'-0" SLD	Fibreglass Waste	I-O
		34'-0" dia. x 16'-0" SLD	Fibreglass Waste	Std.
Oklahoma, Okmulgee — Phillips Petroleum Company		27'-6" dia. x 13'-6" SLD	Oil Refining Waste	Std.
Pennsylvania, Allegheny County — Allegheny Valley Joint Sewage Authority		46'-0" dia. x 16'-0" SLD	Sewage	Std.

Location		Size	Waste	Type
Washington, Ferndale — Mobil Oil Company		50'-0" dia. x 22'-6"SLD	Oil Refining (Phenol)	I-O
	(2)	27'-6" dia. x 14'-0" SLD		Std.
Canada, Elliot Lake, Ontario — City of Elliot Lake		45'-0" dia. x 16'-3" SLD	Sewage	Std.
Cuba, Loma de Tierra — United States Rubber Company, Ltd.		14'-6" dia. x 11'-6" SLD	Sewage	Std.
France, Elbeuf, Seine-Inferieure Ville d'Elbeuf	(2)	65'-6" dia.	Sewage	Std.
Germany, Koln — Esso AG Raffinerie		17'-6" dia.	Oil Refinery Waste	Std.
Germany, Stadt Fuerth — City - Sewage		95'-0" dia.	Sewage	Std.
Japan, Otaru City, Hokkaido City - Sewage	(4)	48'-0" dia. x 16'-0" SLD	Sewage	Std.
Japan, Tokyo, Shibaura Plant City - Sewage	(8)	52'-6" wide x 131'-0" long	Sewage	Std.
Malaya, Singapore — Sembawang Hills Estate		25'-0" dia. x 13'-3" SLD	Sewage	Std.
Mexico, La Galarza, Puebla Compania Ron Bacardi		13'-0' dia. x 9'-6" SLD	Sewage	I-O
Spain, Madrid Alcala de Henares Airport		14'-0"	Sewage	Std.
Sweden, Vintrie Bunkeflo Community		20'-0" dia. x 12'-0" SLD	Sewage	Std.
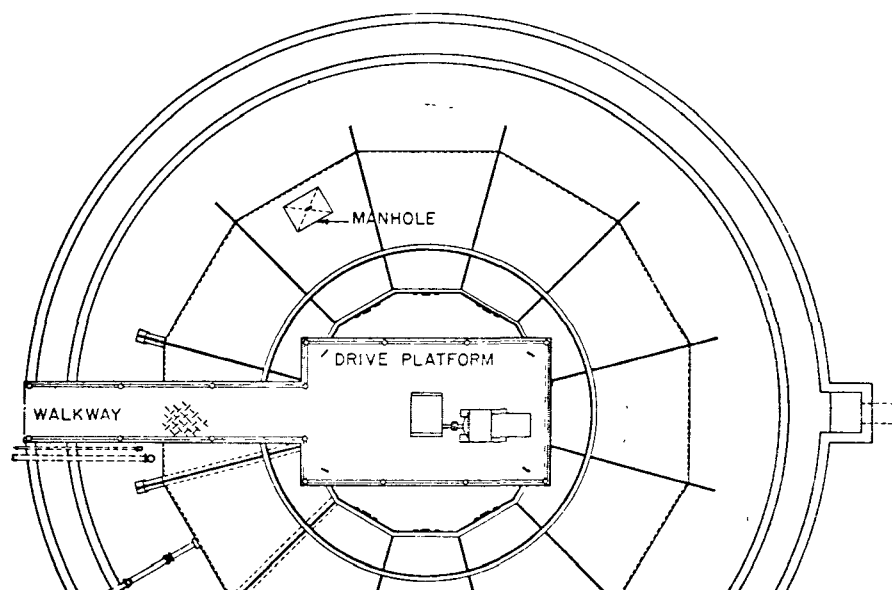
GENERAL

There shall be furnished for installation in each _____(dia, square rectangular), _____ side liquor depth tank with _____ freeboar an "AERO-ACCELATOR" mechanism for complete mixing and biological oxid tion of sewage or waste liquors by intimate contact with aerobic organisms in a activated sludge developed and maintained in an aeration compartment, and fi separation of treated liquid from the activated slurry in a separation compartmer Each unit shall be capable of treating an average flow rate of _____ g.p.r and a peak flow of _____ g.p.m.

"AERO-ACCELATOR" MECHANISM

There shall be furnished all structural and sheet steel to form the inner and out draft tubes, recirculation control gates, hood, support columns or rafters, a bridg consisting of structural beams, checkered plate walkway, handrail and base pla for mounting the drive mechanism at the center of the basin. All steel plates ar shapes shall be furnished knocked down, arranged for field welding by others.

The hood shall form a mixing and aeration compartment for the incoming sewag or waste and activated sludge. The draft tubes shall form recirculation zones an a separation compartment.

AERATOR MECHANISM — Use ONE

USING COMPRESSED AIR

There shall be furnished with each "AERO-ACCELATOR" mechanism a "VORT MIX" Aerator, including a _____ h.p. variable speed drive and reduce couplings, shaft, impeller with straight vertical blades, air sparge ring, support fc sparge ring and all air piping and supports within the basin.

USING ATMOSPHERIC AIR

There shall be furnished with each "AERO-ACCELATOR" mechanism a "VORT AIR" aerator including a _____ horsepower constant/variable speed drive an reducer, couplings, shaft and impeller(s) with straight vertical blades. The aeratc shall be mounted in relation to the surface of the waste and rotated so as to draw large quantities of air from the atmosphere into the liquid.

SLUDGE DISCHARGE

There shall be furnished a waste sludge discharge line consisting of either (a) pipin; of suitable size to pass through the basin wall and a timer operated blow-off valv or (b) an air lift type pump of suitable size including all sludge piping and ai piping within the basin, a timer, solenoid valve and needle valve for the air line and necessary supports.

MOTOR CONTROLS

There shall be furnished an across-the-line starter in NEMA _____ enclosure for_____ phase, _____ cycle, _____ volt current providing over-load and undervoltage protection There shall also be furnished an auxiliary push-button station in NEMA_____enclosure and a push-button station with locking device on the stop lever in NEMA 4 enclosure for mounting at the drive.

BLOWERS required if compressed air and "VORTI-MIX" aerator are used.

There shall be furnished motor drive blowers for supplying air at a pressure of _____ pounds per square inch, as follows:

> _____ unit(s) at_____ cubic feet per minute driven by a _____ horsepower motor.
> One unit at _____ cubic feet per minute driven by a _____ horsepower variable speed drive.
> One unit at_____ cubic feet per minute driven by a _____ horsepower motor for standby service.

There shall be furnished with the blowers, necessary flexible connections, check valves, relief valves, inlet filter silencers, belt guards as required, 0-10 p.s.i. pressure gauge, indicating c.f.m. gauge and suitable orifice plate and flanges. There shall be furnished across-the-line starters in NEMA _____ enclosure for _____ phase, _____ cycle, _____ volt current overload and undervoltage protection and push-button stations in NEMA _____ enclosure.

PAINTING

Before application of paint, all surfaces shall be dry and free of rust or grease. Structural steel shall be given a shop coat of chromic metal primer or equal.

SERVICE ENGINEER

There shall be furnished by the equipment manufacturer the services of an engineer to check the completed installation, place the equipment in operation, and instruct the operators in the correct operation and maintenance procedures.

FULLER
ENGINEERED PRODUCTS FOR INDUSTRY...

AIR HANDLING EQUIPMENT . . . Fuller rotary compressors, Sutorbilt blowers aı Lehigh° fans

PNEUMATIC MATERIALS CONVEYING SYSTEMS . . . Airveyor°, Airslide° aı Fuller-Kinyon conveying systems

POLLUTION CONTROL EQUIPMENT . . . Dracco° dust collectors, Infilco water aı sewage treatment systems

HEAVY PROCESSING MACHINERY AND SYSTEMS . . . Traylor crushers, kilı and mills, Fuller grate coolers, fluid bed reactors and Fuller specialized equipment f the mineral industries

BIOLOGICAL REMOVAL OF PHOSPHORUS

Phosphorus removal by activated sludge treatment is commonly limited to 30-50 per cent by considerations involving bacterial metabolism, the variable quantity of organic matter stabilized by the conventional system and its modifications, and the waste-sludge disposal methods in current use.

Much higher removals are obtained at a few locations and intensive study of these plants is defining the influence of several operational factors on what is termed "luxury uptake" of phosphorus. Among these are the necessity for maintenance of a minimum aeration-basin dissolved oxygen concentration of 2 mg/l, rapid removal of solids from the final clarifier to minimize anaerobic release of phosphorus, and provision of uniform load.

Design of the AERO-ACCELATOR unit is advantageous on all these counts:

1. The oxygen necessary to maintain any desired concentration of mixed-liquor dissolved oxygen is efficiently and economically supplied by its turbine aerator component.

2. Its integral solids separation and recycle systems eliminate the deficiencies of separate final clarification. The activated sludge is constantly being circulated at a high rate from the solids separation zone into the aeration zone.

3. Its complete-mixing design minimizes load variation within the treatment system.

Since 1894...

EXPERIENCED

— offering the broadest, most complete line of equipment for water/waste/reuse systems, ranging from small "package" units to multi-million gallon-per-day treatment plants.

MODERN

— continually adapting to the changing needs of public works and industry. (We believe you will be most interested in the attached small insert, "What has Infilco Done Lately in Solving Pollution Problems?")

CAPABLE

— In working with your problems, with your engineers, our sales engineers have at their disposal all of the resources, technical skills and manufacturing facilities of the General American Transportation Corporation of which Fuller Company – including the Infilco Products group – is a part. Supplementary resources include air pollution control equipment and material handling systems by the Fuller Company; research, testing, development and design through General American Research and Development; and manufacturing plants strategically located throughout the country to allow reduced transportation costs and assure on-time deliveries. The GATX Plate and Welding Division assures united installation responsibility, for customers who desire this service.

NEAR YOU

— As you will wish to initiate action and maintain communication regarding progress, it is important to have service readily available. Sales, service, and application engineers in regional and branch offices stand ready to serve you in the Lake Michigan area.

FULLER
ENGINEERED PRODUCTS FOR INDUSTRY...

AIR HANDLING EQUIPMENT . . . Fuller rotary compressors, Sutorbilt blowers and Lehigh² fans

PNEUMATIC MATERIALS CONVEYING SYSTEMS . . . Airveyor², Airslide® and Fuller-Kinyon conveying systems

POLLUTION CONTROL EQUIPMENT . . . Dracco® dust collectors, Infilco water and sewage treatment systems

HEAVY PROCESSING MACHINERY AND SYSTEMS . . . Traylor crushers, kilns and mills, Fuller grate coolers, fluid bed reactors and Fuller specialized equipment for the mineral industries

Offices in principal cities of the United States and Canada

GATX

ER COMPANY

CO PRODUCTS P.O. BOX 5033, TUCSON, ARIZONA 85703. TELEPHONE 602/623-5401

PHOSPHATE LEVEL CONTROL IN EFFLUENTS

One of the principal factors in the Eutrophication, or aging of lakes, and rivers has been shown to be the phosphate content of waters discharged into the lake or river.

Substantial control of the phosphate content of effluents is possible using . . . .

> ACCOFLOC 6774-C
> ACCOFLOC 6793-C
> ACCOFLOC 6793-D

The ACCOFLOC products are micaceous materials with a platelet molecular structure. The large surface area of such materials is used as a basis for the adsorption, flocculation and removal of phosphate.

CHARACTERISTICS OF ACCOFLOC

ACCOFLOC is mainly inorganic in nature, with organic content of less than 10%. It is non-toxic, disperses rapidly in water, and will give easily pumpable suspensions at solids contents as high as 8% or 80,000 ppm.

A simple agitator will be sufficient to effect dispersion and suspension. The addition of ACCOFLOCS in suspension is recommended for easy control.

The ACCOFLOC flocculates readily in municipal wastes at all points in the processing of the waste, and have been observed to speed the rate of settling of the solids in the waste.

ACCOFLOC PHOSPHATE REMOVAL SYSTEM-AMERICAN COLLOID COMPANY

EFFICIENCY OF REMOVAL OF PHOSPHATE

Suspensions of ACCOFLOC added to Municipal waste removed phosphate as shown on the attached graphs.

PROCEDURE

As the sewage was agitated on a magnetic mixer, the ACCOFLOC was added as a 1% suspension. Mixing was continued for 15 minutes. Sample was removed and analyzed for phosphate.

PHOSPHATE REMOVAL FROM "MIXED LIQUOR"
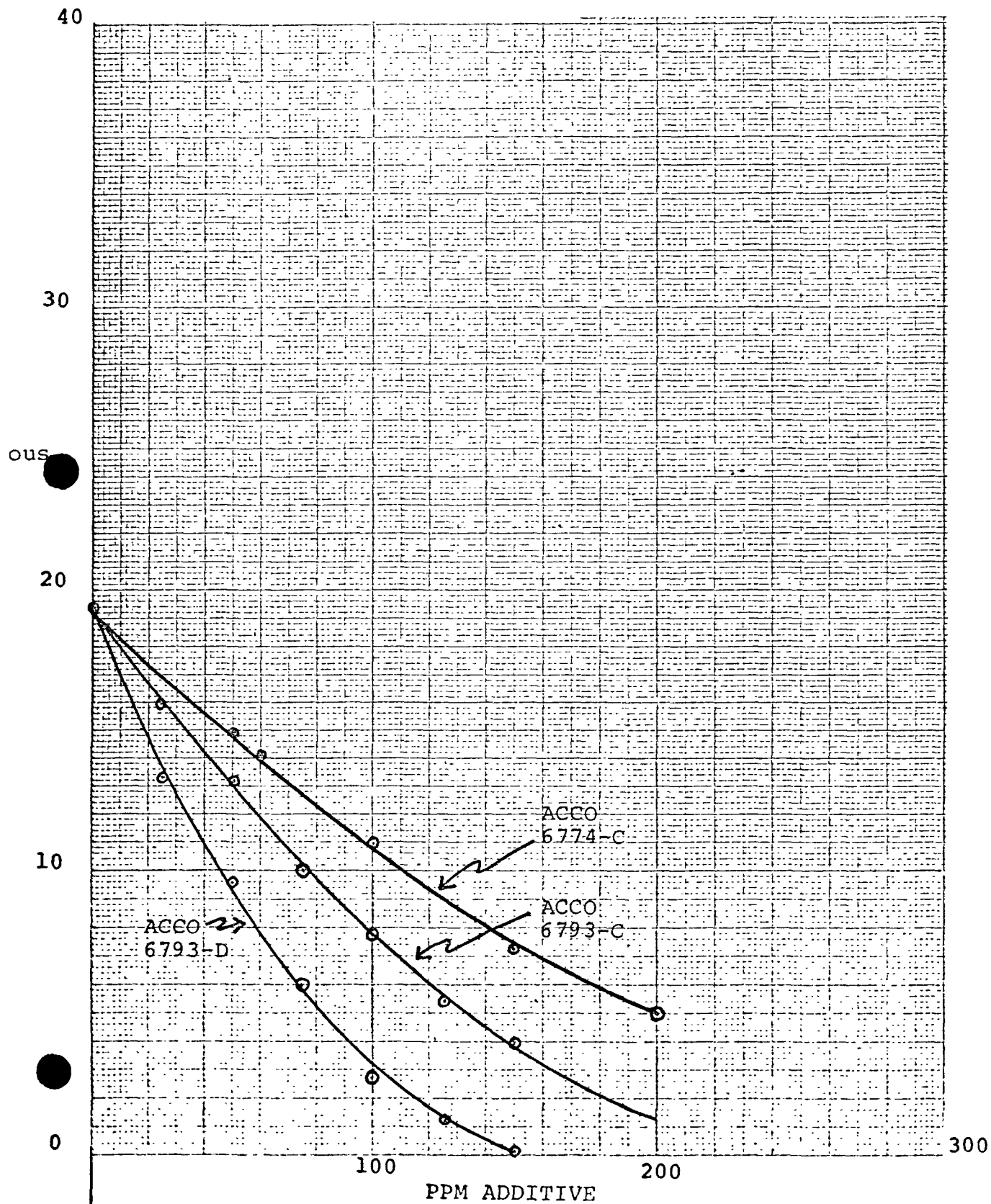
CHICAGO METROPOLITAN SANITARY DISTRICT
HANOVER PARK PLANT

PPM ADDITIVE

ACCO
6774-C

ACCO
6793-C

ACCO
6793-D

ACCOFLOC PHOSPHATE REMOVAL SYSTEM—AMERICAN COLLOID COMPANY

CHICAGO METROPOLITAN SANITARY DISTRICT
HANOVER PARK PLANT

40

30

pm
.e
nor
s
$_4$

20

10

0

ACCO
6774-C

ACCO
6793-D

ACCO
6793-C

100   200   300

PPM ADDITIVE
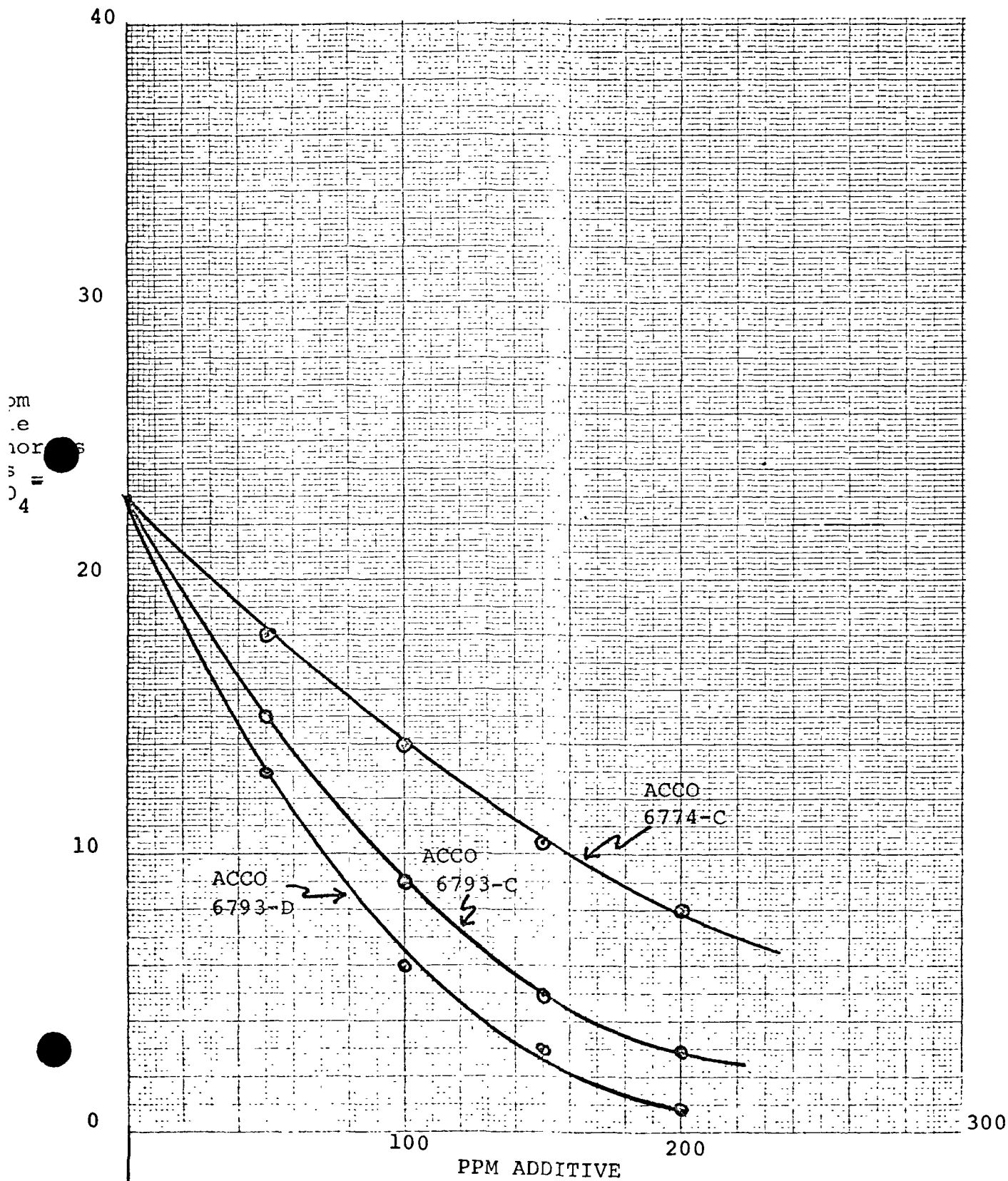
RATE OF REMOVAL OF PHOSPHATE

ACCOFLOC, in suspension, was added to a sewage sample and agitated on a magnetic mixer.

Samples were removed at intervals, as indicated on the graph, vacuum filtered, and analyzed for phosphate.

ACCO 6793-D was used, at a level sufficient to reduce the phosphate content to 20% of its original value, thus effecting 80% removal.

Total Removal = 19.5 - 3.4 = 16.1 ppm

Removal after 2 minutes = 19.5 - 7.5 = 12.0 ppm

Removal after 5 minutes = 19.5 - 3.4 = 16.1 ppm

62.1% of the phosphate was removed after only 2 minutes of reaction. Reaction was complete (82.6% removal) in 5 minutes.

Higher removals, and lower phosphate content effluents could be realized by use of higher levels of ACCOFLOC.

RATE OF REMOVAL OF PHOSPHATE FROM
MIXED LIQUOR

CHICAGO METROPOLITAN SANITARY DISTRICT
HANOVER PARK PLANT

Mixed Liquor With
100 ppm ACCO 6793-D
Suspension Added at
Time Zero

ppm
soluble
phosphorous
$=$
PO_4
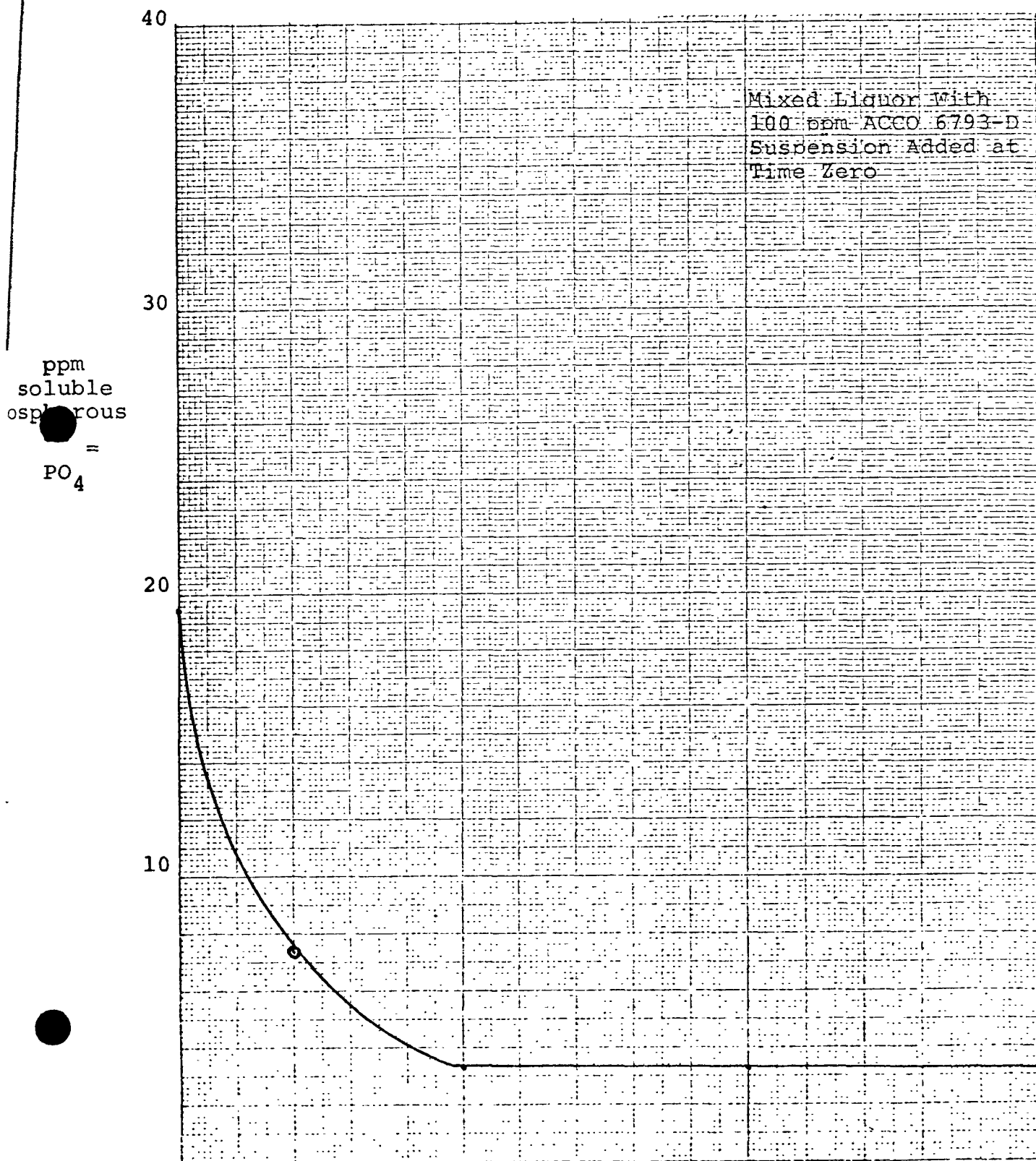
STABILITY OF SETTLED ACCOFLOC

Several biological methods of phosphate removal tend to release the phosphate after settling.

ACCOFLOC was left in contact with the sewage under anaerobic conditions and continuous agitation for extended periods. Samples were taken at regular intervals and a●yzed for phosphate content.

No release or re-solution of phosphate was observed with contact times of up to 30 hours.

STABILITY OF SETTLED PHOSPHATE RICH

ACCOFLOC

CHICAGO METROPOLITAN SANITARY DISTRICT

HANOVER PARK PLANT

MIXED LIQUOR WITH
100 ppm ACCO 6793-D

ppm
soluble
phosphorous
as =
PO_4

40

30

20

10

0

10   20   30
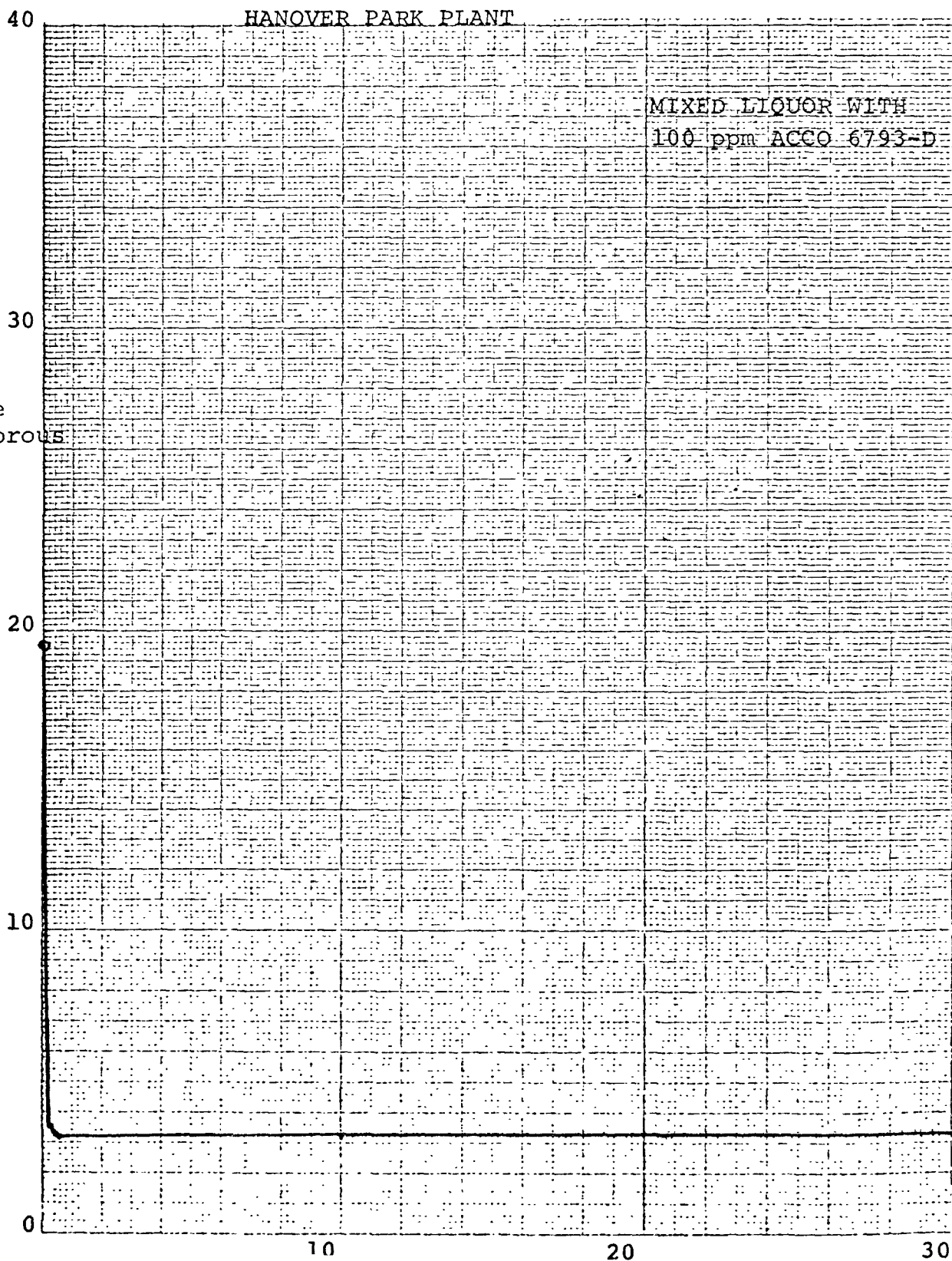
EFFECT ON SETTLING

Laboratory studies indicate that substantial improvement in settling characteristics can be expected by adding ACCOFLOC 6793-D (see following graphs).

With treatment levels of 100 ppm and a low wastage in an activated sludge plant, an equilibrium level of several hundred parts per million would result. As such levels considerable improvement in the settling characteristics can be anticipated.

FREE SETTLING RATES

CHICAGO METROPOLITAN SANITARY DISTRICT

HANOVER PARK PLANT

1000

ACCO-6793-D
added.
Sludge volume of
1000 mls at time
zero minutes

Sludge
volume

500

10 MINS
SETTLING

30 MINS
SETTLING
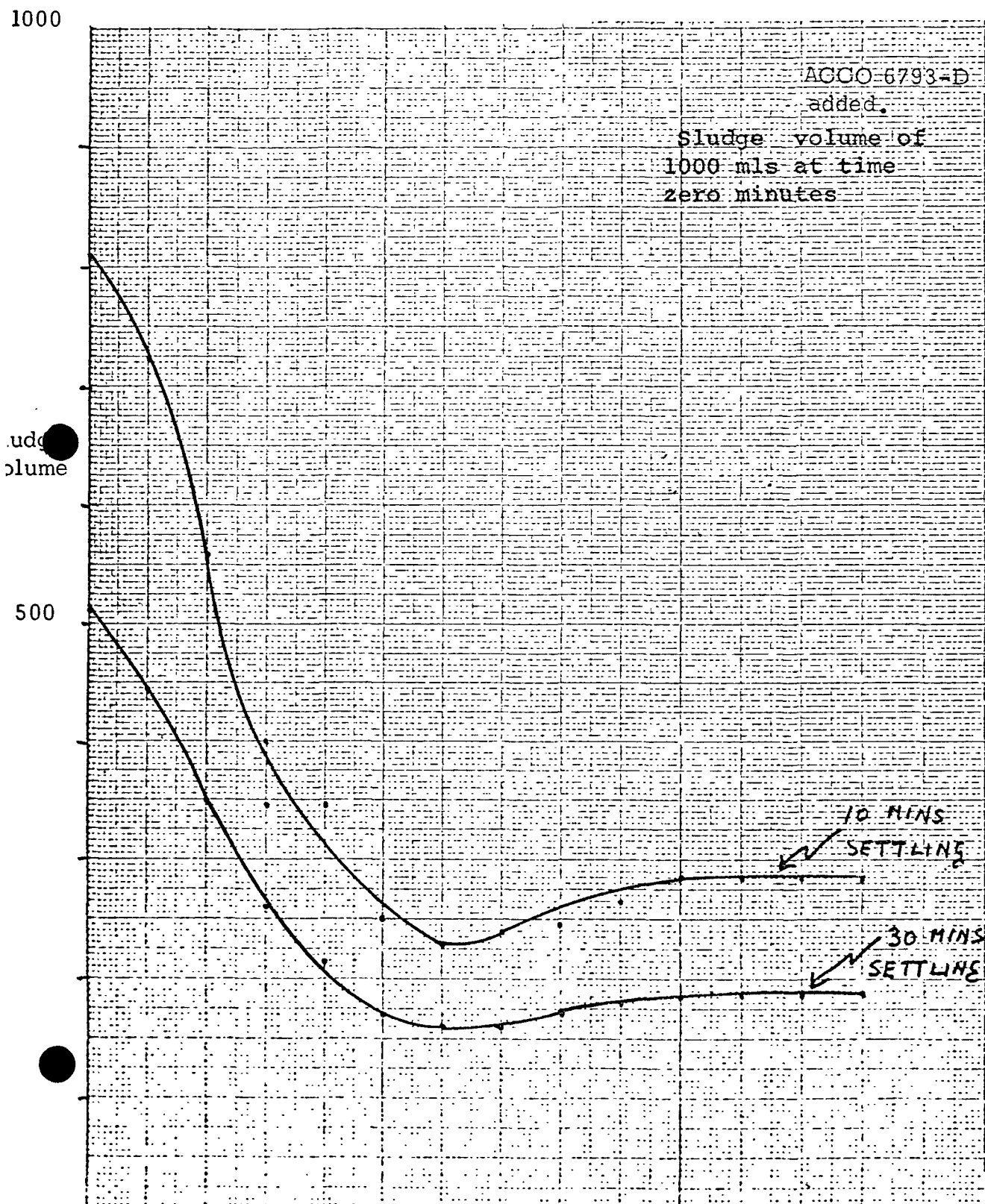
EFFECT ON SLUDGE VOLUME INDEX

CHICAGO METROPOLITAN SANITARY DISTRICT

HANOVER PARK PLANT

ACCO 6793-D
added

300

dge
ıme
ex

200

.00

0   500   1000   1500
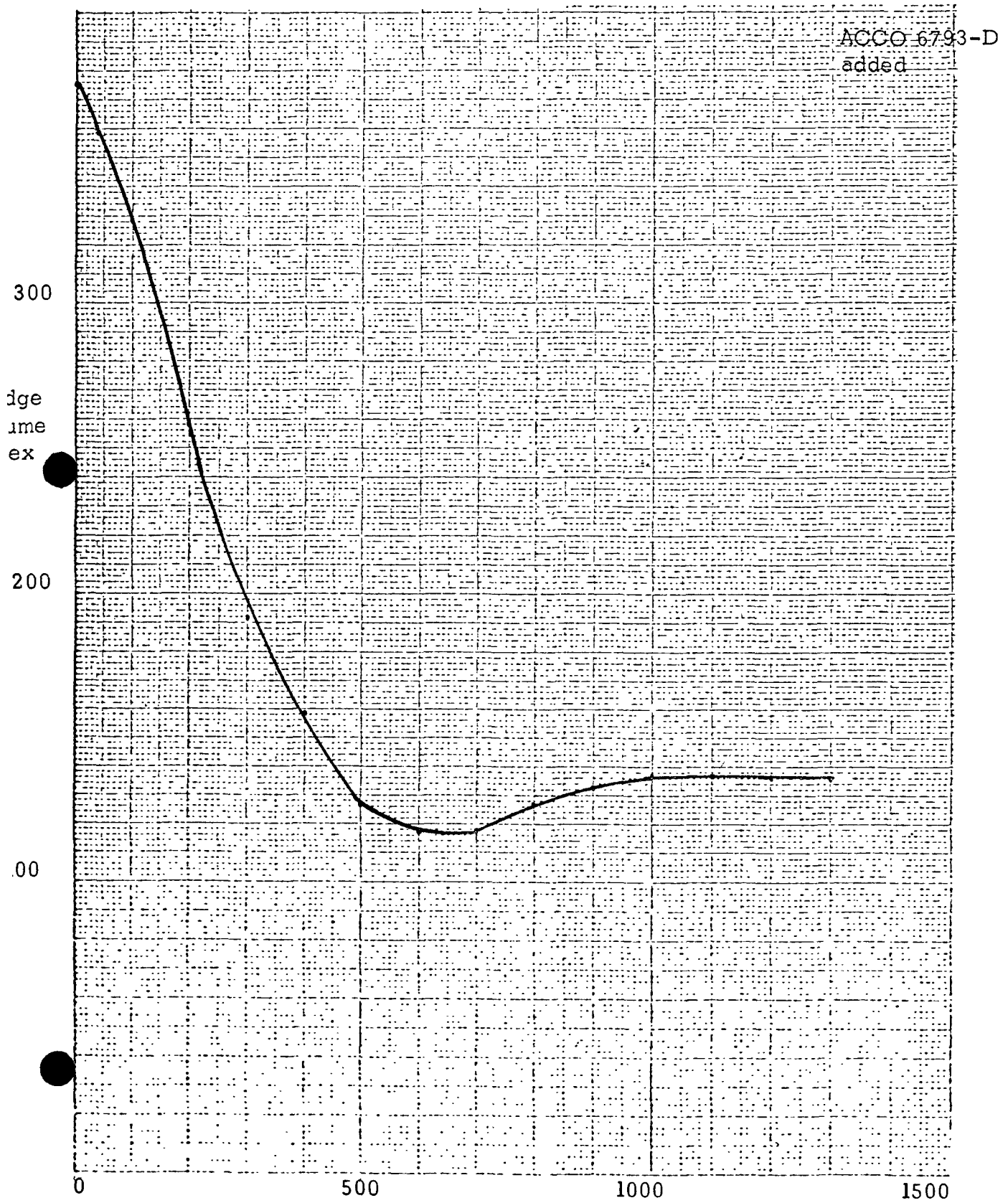
ACCOFLOC PHOSPHATE REMOVAL SYSTEM—AMERICAN COLLOID COMPANY

APPENDIX

PHOSPHATE ANALYSIS

All charts and data show soluble phosphates expressed as ortho-phosphate.

SAMPLE PREPARATION

Sample was filtered through WHATMAN No. 1 paper. 50 mls. of the sample and 2.0 mls of 12 N sulfuric acid were boiled for 30 mins., then cooled to room temperature.

10 mls of molybdate-vanadate solution was added to this sample. Color was allowed to develop for 20 minutes, optical density was measured at 440 milli-microns, and was compared to optical density of known standards of phosphate.

Phosphate content was expressed as parts per million phosphate.

Recla-PaK

From
Raw
Sewage...

to
Polished
Water

complete

- [illegible]
- [illegible]
- [illegible]
- Phosphate removal available as modular accessory
- [illegible]
- [illegible]

Neptune MicroFLOC has already introduced to the waste treatment field the concept of packaged effluent clarifying and polishing equipment through its Model SP Recla-Mate unit (bulletin KL-4560).

Now in Recla-Pak, Neptune Micro-FLOC offers an extremely efficient and economical complete treatment system that will produce *98-99% solids and BOD reduction* to satisfy the most exacting State standards. Biological treatment is combined with efficient tube-type clarification and mixed-media filtration in a single, compact packaged plant.

Recla-Pak provides positive protection against solids being discharged to the receiving stream during any plant upset. The effluent quality of a conventional package plant frequently is degraded by the discharge of large quantities of solids in the plant effluent. These discharges may be caused by surges in the raw sewage flow which overload the conventional package plant settling basin, by mechanical failures of the sludge collection and return system, or by failure of the plant operator to waste adequate quantities of sludge.

Recla-Pak overcomes all of these shortcomings. Extreme fluctuations in raw sewage flow are absorbed in the combination aeration-surge storage tank. This prevents flow surges from upsetting the settling process. *The simple, positive Recla-Pak settling and sludge return system minimizes chances for sludge return failures. The mixed-media filter will trap solids which escape from the settling process.*

Neptune MicroFLOC's advanced research work has been applied to the polishing of sewage plant effluent with the result that more than 98% of raw sewage suspended solids and BOD can now be removed with little increase in operating cost. No chemical coagulants are required. In most cases, clarity of the final effluent exceeds standards of the USPHS for drinking water.

Two MicroFLOC developments are key factors in this process. These are the tube settler which permits the efficient removal of solids within the confines of a relatively small tank and the mixed-media filter which removes remaining solids at high flow rates.

These developments enable communities to meet the most critical standards of sewage treatment with modest capital expenditures, low operating costs, and assurance of returning a high quality effluent to the receiving stream.

More and more water pollution control agencies are setting strict standards limiting the nutrient content of even low solids and BOD effluents. Supplemental equipment is available to treat the effluent either from extended aeration or contact stabilization plants as well as the Recla-Pak to produce effluent phosphate concentrations of less than 0.5 mg/l. Chemical precipitation of the phosphates with alum is followed by tube clarification and mixed-media filtration to produce an effluent low in phosphate. This equipment is available with Recla-Pak as a modular addition to the basic plant.

AUX. SLUDGE WASTE→

AIR COMPRESSOR

AIR

CHLORINE FEED

SCUM

←DECANT RETURN

OVERFLOW

WATER LEVEL

OVERFLOW

EFFLUENT

NORMAL WATER LEVEL

W NAGE

TUBE - TYPE CLARIFIER

MIXED MEDIA FILTER

AERATION TANK

BACKWASH TANK

SLUDGE TANK

FILTER CYCLE ——→   BACKWASH   — — →
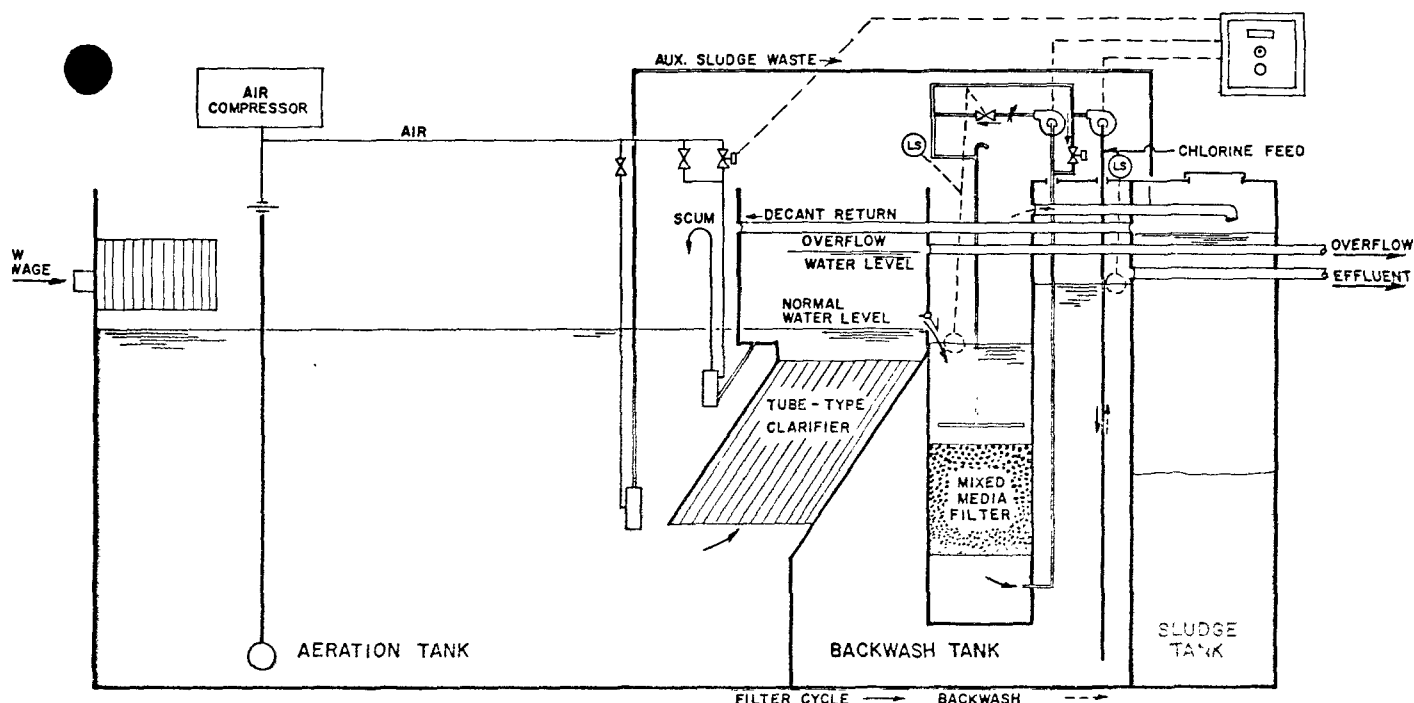
fter ●sing through a bar screen, sewage enters the ᵌratiᵋ ᵇasin where it is subjected to long-term aerobic ological treatment. This biological treatment converts ᵌarly all of the objectionable organic materials in the ιw sewage to biological solids. The aeration period conᵌrts the biological solids formed to a relatively inert ᵓndition, eliminating the need for elaborate sludge gestion facilities.

The aeration tank also serves as a surge storage tank ⌐ allowing the water level to vary so that the settling ᵑd filtration units operate with controlled variations in ᵓw rate.

Water flows from the aeration tank through the unique ιbe settling device. This settling device consists of a ιultiplicity of inclined shallow tubes. The biological ⌐lids formed in the aeration tank are effectively reᵗoved in such a settling unit with only a few minutes' ᵌtention time. The solids are continuously removed ᵒm the settling unit by gravity and returned to the ᵌration process to maintain adequate biological activity ⌐ insure a high degree of biological treatment.

There is no sludge return pump to fail—a weak link the design of most package plants. The very short resiᵗnce time in the settler eliminates the problems caused ⌐ the long settling periods usually associated with packᵌ plants. In other plants, septic conditions can develop the settling tank with the result that sludge will float the surface and be lost in the plant effluent.

Any particulate matter which should escape the setng ●is removed by passing the effluent through a xed-ᵗᵉdia filter. The filter is graded from coarse to ιe in the direction of flow to increase the amount of ιdge which may be stored in the filter, and to provide eater solids removal than would be provided by a plain ᵑd filter. The filter effluent is pumped to the backwash ᵓrage tank which also serves as a chlorine contact tank.

The final plant effluent overflows from this basin.

Recla-Pak effluent is used for backwashing the filter. Backwash is automatic when a pressure sensor detects high headloss across the filter. The backwash water is discharged to the sludge wasting tank and allowed to settle. The supernatant from the sludge wasting tank is then recycled to the aeration chamber. The solids removed from the system by this backwash cycle greatly reduce the frequency with which sludge must be transferred from the aeration basin to the sludge holding tank. During backwash, the incoming raw sewage flow is stored in the combined aeration-surge storage chamber. Scum is transferred to the aeration tank automatically during the backwash cycle.

The backwash storage-chlorine contact tank and sludge wasting tank are integral parts of the compact Recla-Pak structure. All required tankage is contained in a single, compact, rectangular, factory-assembled Recla-Pak unit.

When treating domestic sewage, Recla-Pak consistently provides a final effluent BOD and suspended solids of less than 5 mg/l. The low effluent turbidity (usually less than the drinking water standard of 5 JU) improves the efficiency of effluent chlorination and enables essentially complete removal of coliform bacteria. Recla-Pak combines the best features of biological treatment package plants with all the benefits of effluent filtration. (For improving the quality of existing package plant installations, use Recla-Mate as described in bulletin KL-4560.) Recla-Pak provides a higher degree of treatment and a higher degree of reliability than conventional package plants while decreasing plant space requirements.

GENERAL: Under this section of the specifications, the contractor shall furnish and install a factory-built sewage treatment plant of a capacity of ...... gpd which shall incorporate biological sewage treatment and effluent clarification and filtration. Screening of raw sewage, sludge storage, hydraulic surge storage capacity, chlorine contact time, and filter backwash water storage shall all be part of the package plant. The other principal items of equipment shall include a rotary blower complete with necessary motor controls, air diffusers, filter effluent and backwash pumps, and internal piping. The plant shall be similar and equal to Model -...... manufactured by Neptune MicroFLOC, Inc., 1965 Airport Road, Corvallis, Oregon.

The plant is to provide an average reduction of 98% or more of the BOD and suspended solids normally present in raw domestic sewage. Effluent filtration must be provided as a part of the process to protect against plant upsets and shall function satisfactorily even during plant upsets. Suppliers other than the above-named company wishing to quote on equipment in this section shall submit detail drawings of the proposed equipment and suitable evidence of experience and results to the engineer and shall obtain his written approval to quote at least ten (10) days prior to bid opening.

OPERATION: The plant shall be capable of treating .... gallons per day of raw sanitary sewage with an organic loading of . .... pounds of five-day BOD per day. The maximum design hourly flow shall be .... gph with the maximum flow during a four-hour period being .... gallons.

The raw sewage shall pass through a bar screen to an aeration tank. The liquid level in the aeration tank will be allowed to vary 30 percent from a minimum depth of 7.5 feet to provide storage for hydraulic surges and storage of the influent during the periods when the filter is backwashing. The mixed liquor solids are to be aerated in a clarifier which is an integral part of the aeration chamber. The clarifier shall be arranged so as to provide for continuous gravity return of solids to the aeration tank and is to provide settling surface area of at least one square foot per 100 gallons per day at nominal plant capacity. The clarifier shall provide a Reynolds number of less than 50 at all times to ensure proper hydraulic conditions for sedimentation.

Following the clarifier shall be a filter containing a mixed-media separation bed, graded coarse to fine in the direction of flow so as to optimize sludge storage and to provide optimum solids removal. Flow from the filter shall be to the backwash storage tank which shall provide sufficient storage of filtered effluent for backwashing and for minimum one-hour chlorine contact at rated flow, and shall include means for overflow of the final effluent.

The backwash cycle shall be initiated automatically by headloss across the filter. Manual backwash means also shall be provided. During backwash, the material removed from the filter shall flow to the sludge storage tank. Means shall be provided for transfer of sludge from the aeration tank to the sludge storage tank so that the mixed liquor solids can be readily maintained in the proper range. The supernatant from the sludge storage tank shall be decanted to the aeration chamber.

CONSTRUCTION DETAILS: The package plant shall be factory-built, and of the size and shape shown on the plans. Mechanical simplicity of the plant is deemed important. All components shall be readily accessible for maintenance. The unit shall be designed for installation below grade. All structural shapes shall be structural grade steel not less than $\frac{1}{4}$-inch thick.

The filter shall contain a 30 inch deep mixed-media separation bed composed of three or more materials of different specific gravity and providing a uniformly tapering void distribution from coarse to fine in the direction of flow. Particle sizes shall vary from 0.15mm to 2.0mm.

Internal and external surface preparation and painting, to be performed at the manufacturer's plant, shall include coal tar epoxy to a thickness of 15 mils.

The filter flow rate is to be controlled by a float valve on the effluent side of the filter pump. The float valve is to operate on the level of water above the filter to provide a maximum filter flow rate of 150 percent of the average nominal daily flow rate. The filter pump shall be protected by a float-activated switch.

A control panel shall be furnished and shall include means of automatically programming backwash when filter headloss reaches a preset level, one backwash indicating light, and one pushbutton switch for manual backwash actuation. The control panel shall also include motor starters for blower, effluent pump and backwash pump motors; pressure gauge (headloss); and pressure switch.

The control panel shall be pre-wired in accordance with NEC and CSA standards and shall include a suitable breaker device to receive power input as shown on plans. All exposed electrical devices shall be weather-proof according to NEMA standards. All motors shall be totally enclosed.

The backwash storage tank shall provide water for filter backwashing and shall include a pump suction line. The tank shall be provided with a float-actuated electrical switch with separately adjustable trip and reset points. The switch shall provide pump protection during backwash.

PERFORMANCE GUARANTEE: Contractor shall guarantee that the plant will perform satisfactorily to produce an effluent in accord with the specification standards. The effluent clarifying and polishing system shall be satisfactorily designed to assure that the system will perform continuously within the normal raw sewage loading conditions specified.

EQUIPMENT STARTUP: Contractor shall provide the services of a factory-trained service man for a period of .. days to check out and start the equipment and to instruct the operators in its operation.

Neptune MicroFLOC equipment covered by patents and patents pending.

Recla-Pak—Basic Plant Data

Model Number	Capacity GPD	Total Horsepower	Length* Feet	Weight Thousand lbs
LA-12	12,000	5¼	15	14.0
LA-16	16,000	5¼	20	17.0
LA-20	20,000	7¼	25	18.5
LA-25	25,000	8¾	31	22.0
LA-30	30,000	11	37	26.0
LA-35	35,000	11	43	28.0
LA-40	40,000	13½	49	32.0
LA-45	45,000	14	55	36.0
LA-50	50,000	14	60	38.5

* All Models: Height—10 feet; Width—10 feet

Neptune innovative technology is solving a wide variety of problems in water and wastewater treatment. A nearby Neptune representative has answers.

Advanced Waste Treatment

MicroFLOC filters at Lake Tahoe, California.
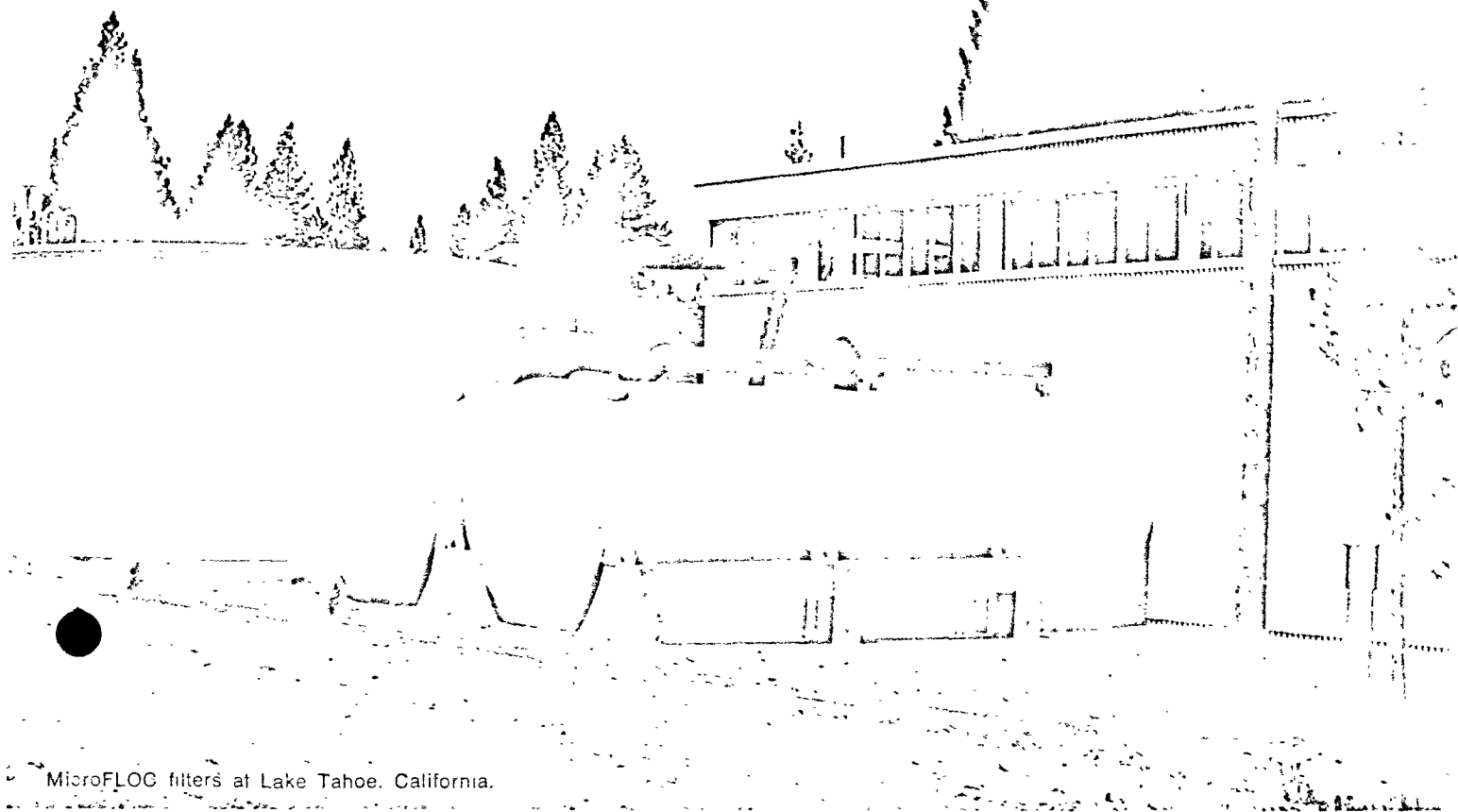
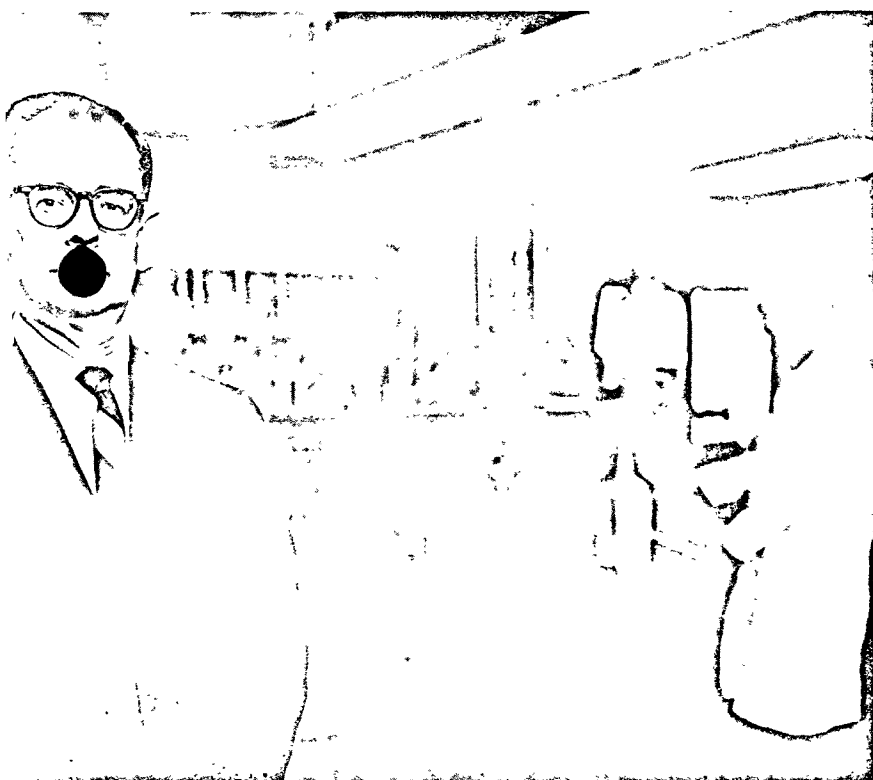# A report on waste water treatment technology and equipment from ...

neptune
FLOC
INCORPORATED®

NEPTUNE MicroFLOC INCORPORATED

A Subsidiary of Neptune Meter Company

The Solution to Pollution... Desire Plus Dollars

In a talk before the 1966 annual meeting of the National Reclamation Association, Mr. Rice said:

"The equation for the solution to pollution is not dilution; it is desire plus dollars ... Pollution abatement is no longer an engineering problem; it is no longer a scientific problem; and it is not a legal problem. It is a problem of politics and economics. If the public will spend the money to do the job, few if any problems cannot be solved.

"During the past few years major changes in the pollution abatement program have resulted from a recognition on the part of the public that the United States economy can afford to solve the pollution problem ... The money and basic technology are available to solve the water pollution problem."

Mr. Rice speaks from a background of more than 25 years' experience in the fields of water and waste treatment—as a sanitary engineer in the Corps of Engineers, as an assistant state sanitary engineer, and as an engineer specializing in the design of water and waste treatment equipment. He is the 1966 recipient of a George Warren Fuller Memorial Award from the American Water Works Association.

Archie H. Rice, president of Neptune MicroFLOC, Inc., in the company's Corvallis research and development laboratories.

Neptune MicroFLOC *Innovations Contribute to Advanced Waste Treatment Technology*

Application of scientific theory to the development of practical processes has resulted in several contributions by Neptune MicroFLOC to basic technology of advanced treatment of domestic wastes and industrial effluents. Innovations in areas of solids separation, filtration and process control have been made and demonstrated under practical operating conditions. New equipment has been developed utilizing these technological advances.

Efficient Solids Separation in Less Than 10 Minutes

Several years of Neptune MicroFLOC research and development work have resulted in a unique, proven application of basic settling theory. Excellent sedimentation is achieved with less than 10 minutes detention time, compared with several hours required in normal settling basins.

The compact solids separation device is packed with long, shallow tubes. The flow is passed lengthwise through these tubes, each acting as a shallow settling device. Settleable material is deposited on the bottom of each tube and the clarified effluent is continuously discharged. This tube settler is inclined upward in the direction of flow to provide for gravity drainage of the solids. Settling particles need fall only a fraction of an inch in the tubes rather than many feet required in large settling basins.

This settling device is in operation in several Neptune MicroFLOC waste and water treatment facilities in North and South America, and is providing excellent results while saving space and capital costs.
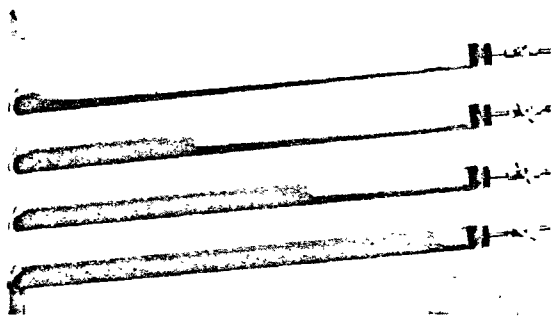
This photograph demonstrates how the tubes progressively fill with floc. The bottom tube is almost full and ready for backflushing. The other tubes were placed in service at later intervals to show the progressive advance of the "rolling front" of sludge deposits.

High-rate Filtration of Sewage Effluent

Another innovation in advanced waste treatment is high-rate filtration. MicroFLOC mixed media filter beds are graded from specially selected coarse media at the top to very fine media at the bottom. This system in effect provides decreasing void sizes in the direction of flow and results in exceptional stability, high filtration efficiency and ability to store large quantities of material between backwashes.

Use of coarse material avoids plugging at the filter surface, a common problem in attempts to apply sand filters to waste treatment. Similarly, the top-to-bottom gradation enables much finer filter material to be used. Media as fine as 0.15 mm can be used in these high-rate filters. MicroFLOC mixed media filters are producing filtrate of high clarity from secondary sewage effluent, operating at 5 gpm per square foot.

Automatic Controls Improve Efficiency, Reduce Costs

MicroFLOC has extensive experience in the development and manufacture of automatic control systems to minimize process malfunctions due to errors or inattention, and to reduce operating costs. Automatic MicroFLOC water and waste treatment plants utilizing these control systems are giving excellent results.

New Approaches to Industrial Effluent Treatment Problems

MicroFLOC research and development people are constantly working with industry to examine new approaches to industrial effluent treatment problems. Some of the pilot studies made to examine feasibility of industrial effluent treatment include the following:

Separation of cereal grains from hexane stream.

Tertiary treatment of meat packing effluent.

Treatment of process waste waters from electronics manufacturing.

Treatment of effluent from preservative processing of piling.

Treatment of plywood mill (glue) effluent.

Treatment of wastes from hydraulic gravel classification.

Treatment of milk processing effluent.

Treatment of steel mill effluent.

Tertiary treatment of combined municipal-cannery wastes.

Treatment of metal plating effluent.

Treatment of effluent from gypsum wall board manufacturing.

Soft drink rinse water reclamation.

Reclamation of laundry waste water.

Clarification of flue gas scrubber stream.

Degree of Treatment *Can Be Tailored To Fit Effluent Quality Required For Pollution Control or Reuse*

Increasing concern over pollution of the nation's rivers and lakes is resulting in higher standards of water quality at national, regional and state levels. The traditional approach to waste disposal has been to provide minimum treatment and depend on the natural purification capacity of the receiving stream to do the rest of the job. This use of waterways for waste disposal is in head-on conflict with the growing demands of an affluent society for recreational uses and for preservation of scenic values of our waters.

As a result, the trend now is for subjecting waste water to higher degrees of treatment, thereby minimizing dependence on natural processes. Conventional sewage treatment processes do not remove many waste water constituents which are of increasing concern. For example, sewage which has been treated by conventional means still contains material which will stimulate growths of algae in the receiving body of water. These algal growths can interfere with recreation and downstream use.

Many authorities agree that the time is near when the degree of waste water treatment required in many areas will be so costly that cities and industry cannot afford the luxury of discarding water after only one use. In many areas of the world, it is only a matter of time before serious consideration must be given to direct reclamation and reuse of waste water to supplement inadequate potable supplies. Already, complete utilization of available potable water supplies has forced one South African city to the direct potable use of reclaimed waste water. Within the United States, reclaimed waste waters are being used for irrigation, industrial use, groundwater recharge, and recreational lakes. However, reclaimed wastewater is still a generally untapped water resource which offers an economical solution to many water problems.

Technology and equipment now are available to provide the degree of treatment of waste water necessary to produce any quality of reclaimed finished water required. The table below shows some examples of how effluent quality can be tailored to meet requirements. Reclaimed water of high clarity and suitable for many industrial purposes can be produced with simple mixed-media filtration. Chemical coagulation, filtration and adsorption on activated carbon can be employed for higher quality effluent, with removal of phosphates, color and odor.

Qualities of Effluent by Various Processes

PROCESS	SUSPENDED SOLIDS (mg/l)	BOD (mg/l)	COD (mg/l)	TURBIDITY (mg/l)	COLOR (Units)	PO$_4$ (mg/l)
TYPICAL EFFLUENT QUALITY						
Simple Mixed-Media Filtration of Activated Sludge Effluent	1 to 5	1 to 5	40 to 70	0.3 to 5	15 to 50	20 to 40
plus Activated Carbon	1 to 3	<1	5 to 15	0.3 to 3	<5	20 to 40
Coagulation and Mixed-Media Filtration of Secondary Effluent	<1	<1	20 to 60	0.1 to 1.0	10 to 30	0.1 to 1.0
plus Activated Carbon	<1	<1	1 to 15	0.1 to 1.0	<5	0.1 to 1.0

Recla-Mate SP Provides 99% Overall BOD Reduction for Only 10% Increase in Operating Cost

Operating plants demonstrate that excellent results can be achieved in simple mixed-media filtration (no chemicals used) of extended aeration effluents. Operating at filter rates of 5 gpm per square foot, overall reductions of 99 percent of the raw sewage BOD and 98 percent of the raw sewage suspended solids are provided by simple mixed-media filtration of extended aeration effluent. The unique Recla-Mate process produces filtrate of high clarity even during severe upsets of the extended aeration plant.

Skid-mounted equipment (Recla-Mate, Series SP), factory assembled and tested, is available for providing this degree of treatment for smaller package plants. The same technology is applicable to large municipal and industrial plants using field erected equipment.

Recla-Mate unit is achieving excellent results with extended aeration effluent at U.S. Forest Service Job Corps Center at Camp Angeil, Oregon.
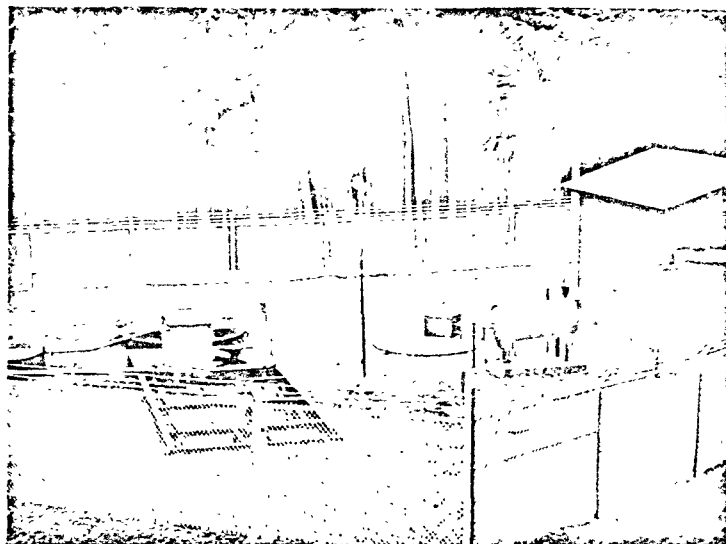
Recla-Mate SWB Achieves High Degrees of Turbidity and Nutrient Removal

The use of Recla-Mate SWB with coagulants provides a means of reclaiming water suitable for many uses. It is a means of preparing secondary effluents for activated carbon treatment and demineralization. Efficient coagulation and filtration of secondary effluent produces a final effluent with suspended solids and BOD of less than 1 part per million.

Phosphates can be reduced to less than 1 part per million by using sufficient coagulant. Soluble organic compounds which contribute to the color of the filtered effluent can be effectively removed by adsorption on activated carbon.

Factory assembled and tested Recla-Mate is available to provide flocculation, sedimentation and mixed-media filtration up to 140,000 gpd in a single compact unit. Such a unit is in use with MicroFLOC ion exchange and activated carbon units at Ely, Minnesota.

The Federal Water Pollution Control Administration recently installed a 20,000 gpd unit at Ely, Minnesota. It is similar to the 140,000 gpd unit shown.
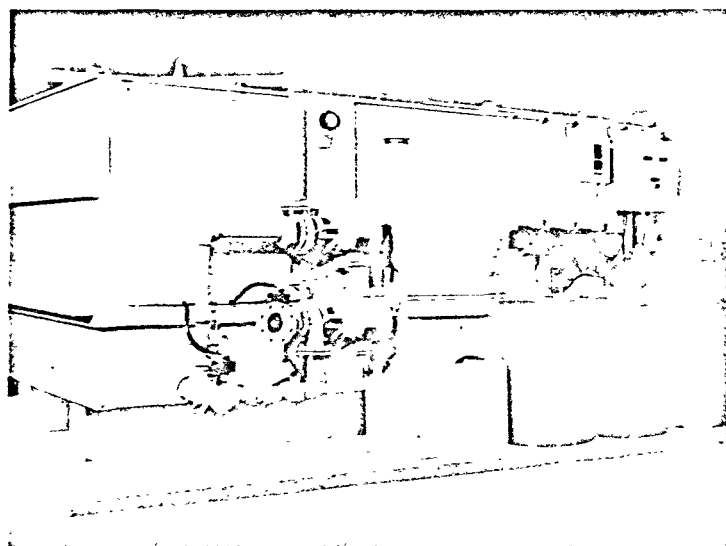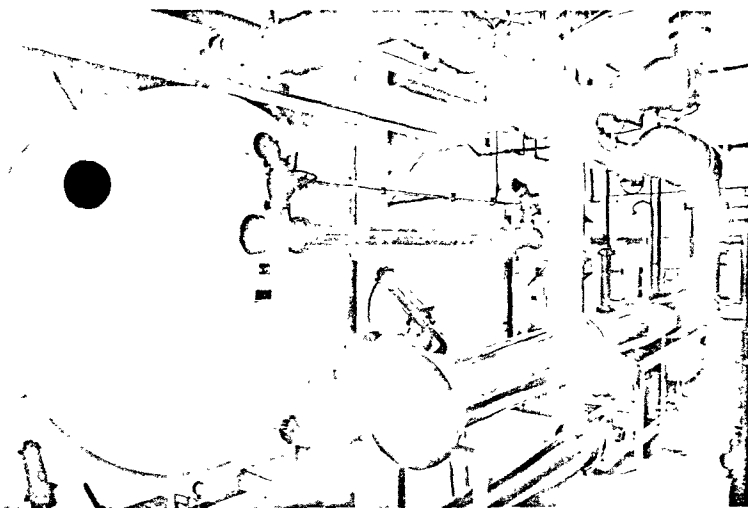
Coagulated secondary effluent is filtered on MicroFLOC mixed-media filters in waste water reclamation plant at Lake Tahoe, California.
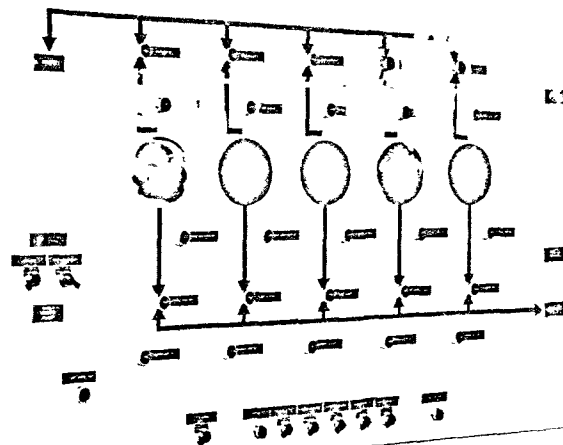
MicroFLOC monitoring and control systems provide automatic operation of advanced waste water treatment plants

Process Applied to Large Automatic Plants For Municipal, Industrial Use

A 2.5 MGD water reclamation plant using a patented MicroFLOC advanced waste water treatment process is in operation at Lake Tahoe, California. Coagulated secondary effluent is filtered by MicroFLOC mixed-media. At the Tahoe plant, filtered effluent has a BOD of less than 1 part per million, a phosphate content of 0.1 to 1.0 parts per million, a color of 10 to 30 units, a turbidity of 0.1 to 1.0, a coliform content of less than 2.1/100 ml following chlorination, and contains no detectable virus.

This high clarity filtered effluent then receives treatment on granular activated carbon. The carbon columns produce a colorless, odorless effluent, free of virus and coliform bacteria. Spent carbon is dewatered and thermally regenerated at the Tahoe plant.

Cost of regenerated carbon averages about one-sixth the cost of virgin carbon. Coagulant is recovered and reused to further reduce operating costs.

Additional MicroFLOC filters are being installed at the Tahoe plant to bring capacity up to 7.5 MGD. At the same time one more degree of treatment will be provided with the addition of a stripping tower to remove 95 per cent of the ammonia nitrogen. Use of lime as the coagulant raises the pH to a favorable range for ammonia stripping at no additional chemical costs.

Costs Range from $10 to $150 per Million Gallons

Water reclamation is an economically available, dependable and easily obtainable water resource as well as a positive means of pollution control. The degree of treatment can be tailored to specific effluent quality requirements. Costs are proportionate to the degree of treatment. For a 10 MGD plant, costs (including capital, operating, maintenance costs) will range from $10 per million gallons for plain filtration of secondary effluent on mixed-media filters, to $150 per million gallons for chemical coagulation, sedimentation, filtration, activated carbon and ammonia stripping.

Why Neptune MicroFLOC Leads *in* Advanced Waste Treatment Technology

The leadership in technology and equipment for advanced treatment of domestic waste and industrial effluent which Neptune MicroFLOC has acquired reflects the capabilities of the company's staff of process specialists. They are "state of the art" men who have made original contributions to the field and are responsible for many of the developments in waste water reclamation. The process experts are supported by a large staff of experienced mechanical, chemical and electrical designers.

Walter R. Conley
Director of Research and Technical Services. A chemical engineer who has pioneered and developed many of today's advanced water treatment concepts.

Gordon L. Culp
Research Manager, B.S , Civil Engineering. M S., Sanitary Engineering; author of many published papers on waste water reclamation; has extensive experience in waste water reclamation research and plant scale application of advanced treatment processes

Sigurd P. Hansen
Research Engineer, B S , Civil Engineering, M. S., Sanitary Engineering; experienced in advanced treatment of industrial and municipal wastes, author of several published papers.

John R. Stukenberg
Research Engineer, B.S., Civil Engineering, M.S ,PhD , Sanitary Engineering; specializing in biological and advanced treatment of industrial and municipal wastes

Selected Published Papers

Hansen, S. P., and *Culp, G. L.*, "How to Clean Waste-water for Reuse." *American City,* (June, 1967).

Conley, W. R., and Evers, R. H., "Coagulation Control." Presented at 1967 AWWA Conference, Atlantic City, N.J. *(June, 1967).*

Culp, G. L., and Hansen, S.P., "Reclamation of Waste Water for Reuse." Accepted for the International Conference on Water for Peace, Washington, D.C., May, 1967.

Slechta, A. F., and Culp, G. L., "Water Reclamation Studies at the South Tahoe Public Utility District." *Water Pollution Control Federation Journal,* (May, 1967).

Culp, G. L., and Hansen, S.P., "Extended Aeration Effluent Polishing by Mixed-Media Filtration." *Water and Sewage Works* Vol. 114, pp. 46-51 (February, 1967).

Miehe, F. J., "High-rate Filtration of Process Water." *Pulp and Paper Magazine of Canada,* p. 90 (Jan. 1967).

Slechta, A. F., and Culp, G. L., "Phosphorus and Nitrogen Removal at the South Tahoe P.U.D. Water Reclamation Plant." Presented at 39th Annual Conference *of the Water Pollution Control Federation,* Kansas City, Missouri (September 29, 1966).

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Stukenberg, J., "Water Pollution and Biological Treatment." *Kansas Engineer,* p. 14 (April, 1966).

Culp, G. L., and Slechta, A. F., "Tertiary Treatment Practice Studies of Carbon Adsorption, Coagulant Recovery, and Nutrient Removal at Lake Tahoe." Presented at 38th Annual Conference of the California Water Pollution Control Association, Monterey, California (April 28, 1966).

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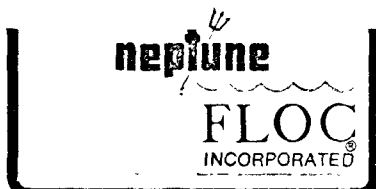
nepjune presents

Complete Package Sewage Plants and Supplementary Upgrading Units
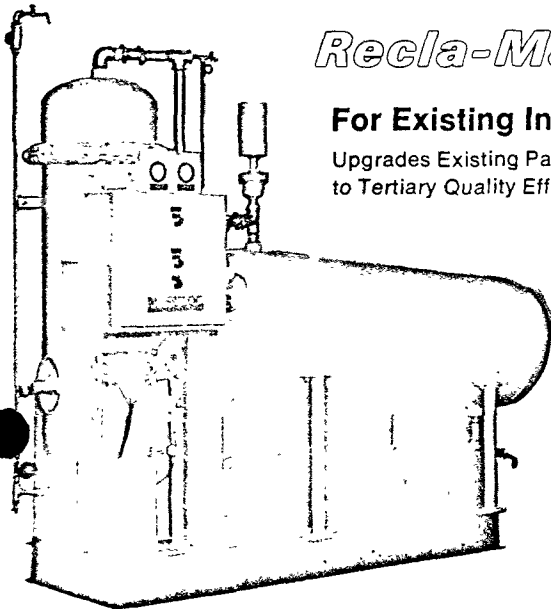
Recla-Mate SP

For Existing Installations

Upgrades Existing Package Plants
to Tertiary Quality Effluent

Recla-PaK

A Completely Self-Contained Unit

Treats Raw Sewage to
Tertiary Quality
Effluent

98% Solids Removal
99% BOD Removal
Positive Control of Effluent
 Quality
Automatic Operation
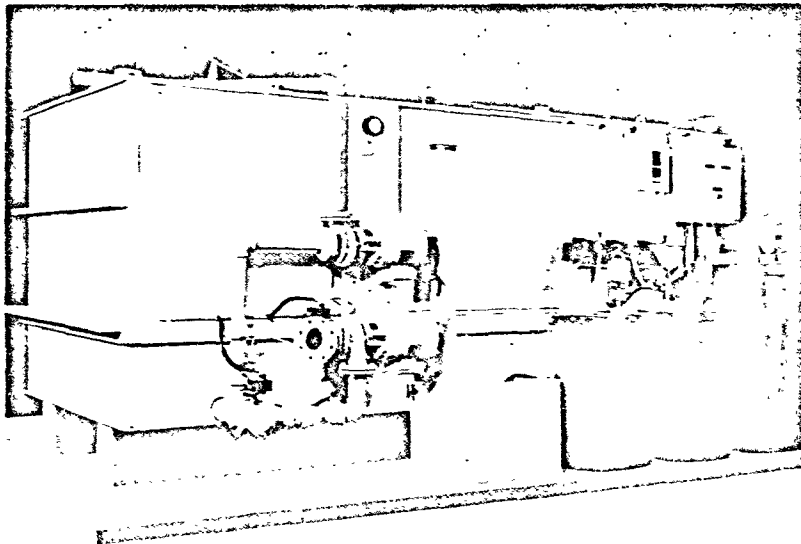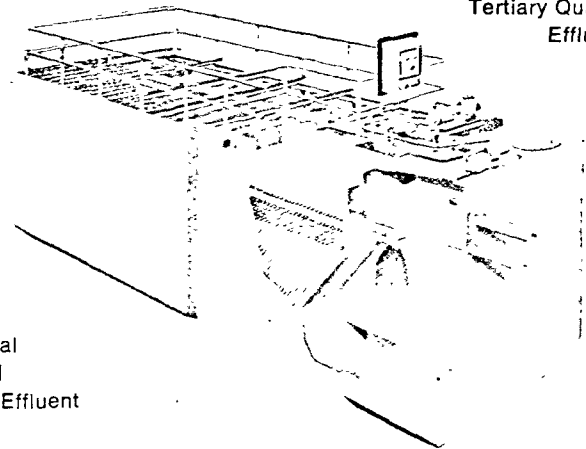
Recla-Mate SWB

**Phosphate Removal
for Existing Plants or
New Installations**

Phosphate concentrations
reduced to less than 0.5 mg/L

Clear Water NOW through Years-Ahead Technology

neptune FLOC

WATER AND WASTEWATER TREATMENT DIVISION OF
NEPTUNE METER COMPANY

Recla-Pak — The complete sewage treatment plant. (Tertiary quality from raw sewage.) Recla-Pak offers an extremely efficient and economical sewage treatment system in one package that produces effluent quality of 98 to 99% solids and BOD reduction from raw sewage. Biological treatment is combined with efficient tube-type clarification followed by mixed-media filtration. Recla-Pak's unique clarification/filtration features provide positive protection against solids being discharged to the receiving stream. This unit satisfies the most demanding pollution control standards.

Recla-Mate "SP" — Tertiary treatment of effluent from existing package sewage treatment plant installations. Designed to "polish" the effluent from biological package sewage treatment plants, Recla-Mate employs tube-type clarification and mixed-media filtration to give overall reductions of 98% solids and 99% BOD. Recla-Mate offers these features: a minimum of maintenance and operator attention and continuous production of high quality effluent, even during upsets of the existing upstream plant. Recla-Mate was designed specifically to upgrade existing biological package plants to meet rigid pollution control standards.

Recla-Mate "SWB" — Phosphate reduction for package sewage treatment plant installations. Recla-Mate SWB is an automatic package unit which employs coagulation, flocculation, tube-type clarification, and high rate, mixed-media filtration. Recla-Mate SWB makes feasible the tertiary treatment steps required to meet those pollution control standards demanding high degrees of phosphate removal.

Recla-Mate and Recla-Pak units are easily transportable to the jobsite. Installation typically consists of pad or burial site preparation and completion of field-ready piping and wiring connections. The automatic nature of these package tertiary plants means low time, cost, and skill demands for operation.

Send for specific application data

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Send for specific application data

THE EIMCO CORPORATION

WATER TREATMENT EQUIPMENT

...For Industrial and Municipal Applications
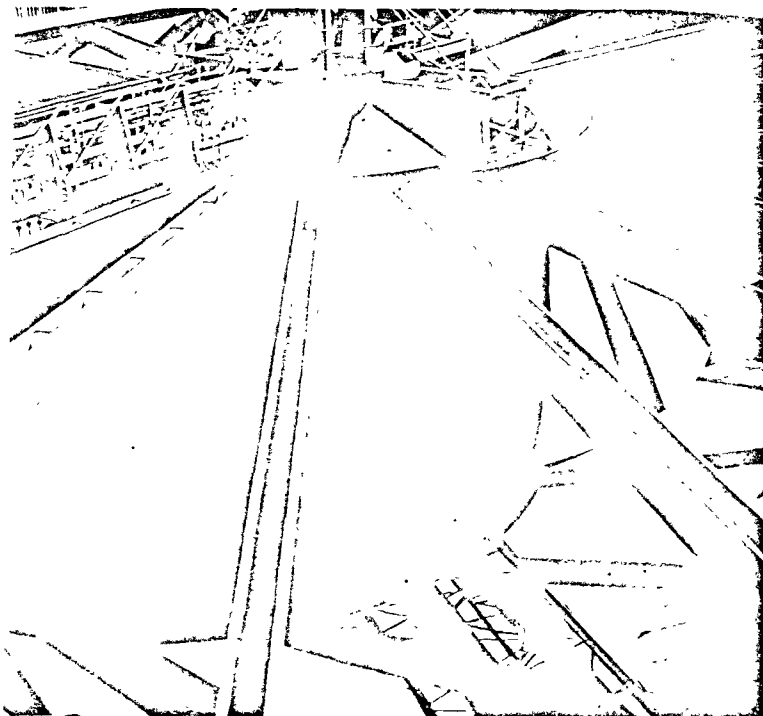
Eimco Reactor-Clarifier™ water treatment unit is a highly versatile machine that combines flocculation, coagulation, clarification and positive sludge removal in a single tank. For municipal or industrial use, these are the most compact and economical to operate units available today. They will remove turbidity, algae, color, iron and other contaminants. They accomplish lime or lime-soda softening, magnesium precipitation, brine softening or clarification and wastewater clarification.
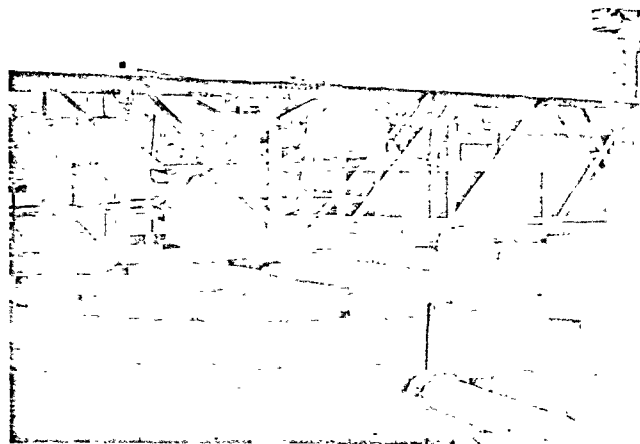
Solids-Contact Types HRC and HRB . . . use th proven, highly efficient upflow solids contact action. Large diameter turbines internally recirculate large quantities of previously forme floc or precipitates at low peripheral turbine speeds. In softening operations, this recirculatic can be up to 15 times the feed rate with slurry density up to 5 per cent by weight. Excellent overflow qualities and dense underflows are obtained in this simple, stable operation. The is no unstable sludge blanket to upset the operation.
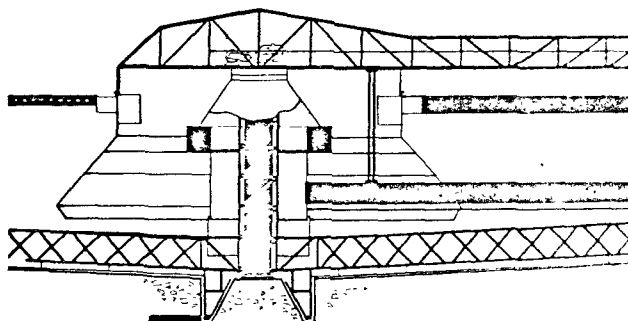
Type HRC from 50 to 200 ft. diamet
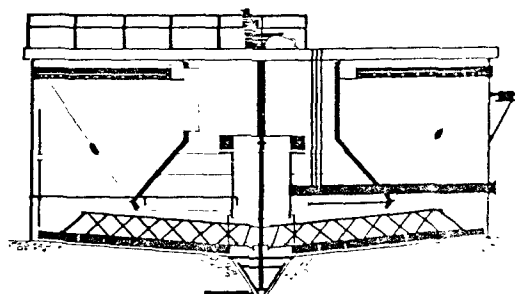
Type HRB from 10 to 75 ft. diame'

Reactor-Clarifier™
treatment units
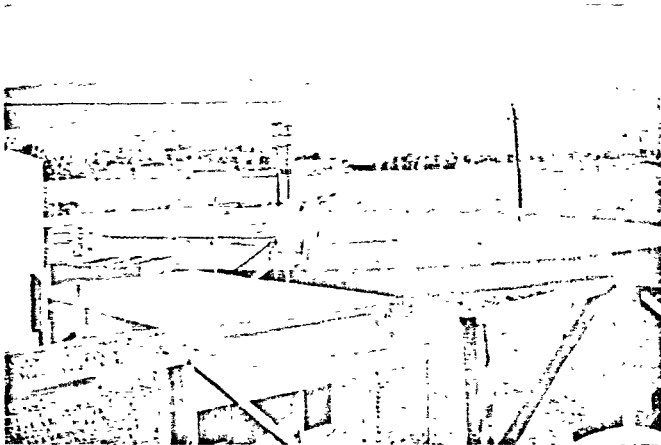
Type CF — Standard rate 30 to 200 ft. diameter, center column supported. This is a standard rate unit which combines vertical paddle flocculation with clarification. Center column units in sizes from 30 to 200 ft. diameter are standard. Recommended for turbid water clarification, algae and color removal. The standard unit has a uniform influent distribution system.
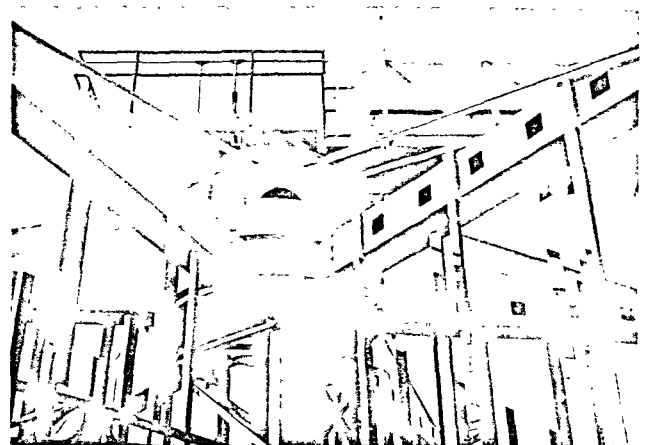
Type BFR — Standard rate type up to 70 ft. diameter, beam supported. The BFR is a standard rate type that combines slow speed turbine flocculation with clarification. Beam supported units only for diameters 20 to 70 ft. For turbid water clarification and treatment of industrial wastes where gentle flocculation by turnover is beneficial.
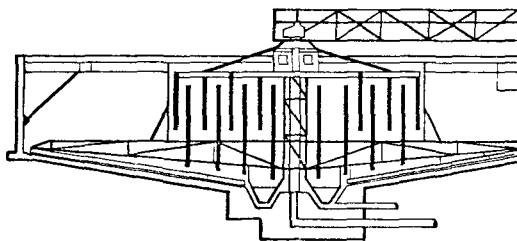
Type CF from 30 to 200 ft. Diameter
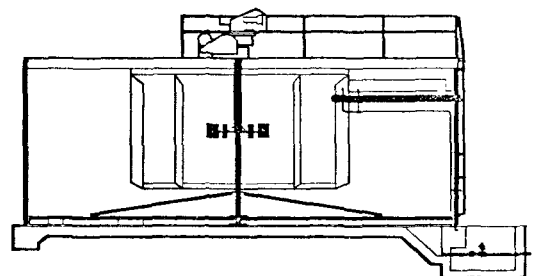
Type BFR from 20 to 70 ft. Diameter

Typical industrial users of Eimco
Reactor-Clarifier treatment units.

Upper left

The Southern Nevada Power Company's Sunrise
Station Effluent Treatment Plant includes a
60-ft. diameter Reactor-Clarifier for solids
contact cold lime treatment to remove phosphate
from sewage treatment plant effluent prior to
cooling use. The sewage effluent contains
from 15 to 35 ppm as Ortho-PO_4 and less than
1 ppm after treatment. — *Stearns-Roger
Corp. engineers*

Left center

U. S. Steel Corporation's plant near Provo, Utah
utilizes Reactor-Clarifier treatment units for
mill scale water treatment. Special skimming
devices are used to remove oil.

Lower left

Two 55-ft. diameter high rate Reactor-Clarifier
units at Kaiser Steel Corporation, Fontana,
ornia. Six Reactor-Clarifier units are used at
this plant for treating various types of steel
mill waste water treatment and for
water reuse. — *Kaiser Engineers*

Upper right

Process water for the Allied Paper Company
kraft mill at Jackson, Alabama is treated in a
150-ft. diameter by 32-ft. sidewall depth
Reactor-Clarifier treatment unit. High color and
turbidity removal of river water is accomplished
at rates up to 26 mgd. In addition to clarification,
the unit provides storage of 1 million
gallons of water. The launders are
submerged eight feet below surface. — *Eastern
Engineering Company*

Lower right

A high rate Reactor-Clarifier treatment unit at
Northwest Paper Company's mill at Cloquet,
Minnesota, removes turbidity from 30 to
40 mgd of river water.

Municipal users of Eimco
Reactor-Clarifier treatment units include
both large and small plants.

Upper left
The residents of Merida, Yucatan in Mexico now
have a highly efficient treatment plant. Two
90-ft. diameter Eimco Reactor-Clarifier treatment
units form the basis for the high rate softening
plant. — *Charles S. McCandless Co.,*
engineers

Lower left
At Titusville, Florida, the high rate Reactor-
Clarifier treatment unit, 50-ft. diameter, is used
for turbidity removal and softening. The heavy
duty construction of the unit makes it possible
to handle the heavy sludge which results from
softening. Plant capacity is 6 mgd. —
Black, Crow and Eidsness, Inc., engineers.

Upper right
The resort city of Aspen, Colorado has a 4 mgd
tment plant which uses a type CF
ctor-Clarifier treatment unit,
75-ft. diameter. — *Dale H. Rea, engineers*

Center right
Four high rate Reactor-Clarifier water treatment
units are installed at the Rinconada Water
Treatment Plant of the Santa Clara County,
California, Flood Control and Water
District. — *Kennedy Engineers*

Lower right
At State University of Iowa, two Reactor-Clarifier
treatment units are used in the plant for both
the city and university. Students use the plant
for laboratory and research work. — *Stanley*
Engineering Company

Upper photo
Water treatment plant for the fine paper company of P. H. Glatfelter
Co., Spring Grove, Pennsylvania, includes this 105-ft. diameter
Reactor-Clarifier treatment unit, type HRC. — *J. E. Sirrine Co., engineers.*

The San Geronimo plant of Marin Municipal Water District near
San Rafael, California uses this 125-ft. diameter Reactor-Clarifier unit in
the treatment process. — *Kennedy Engineers.*
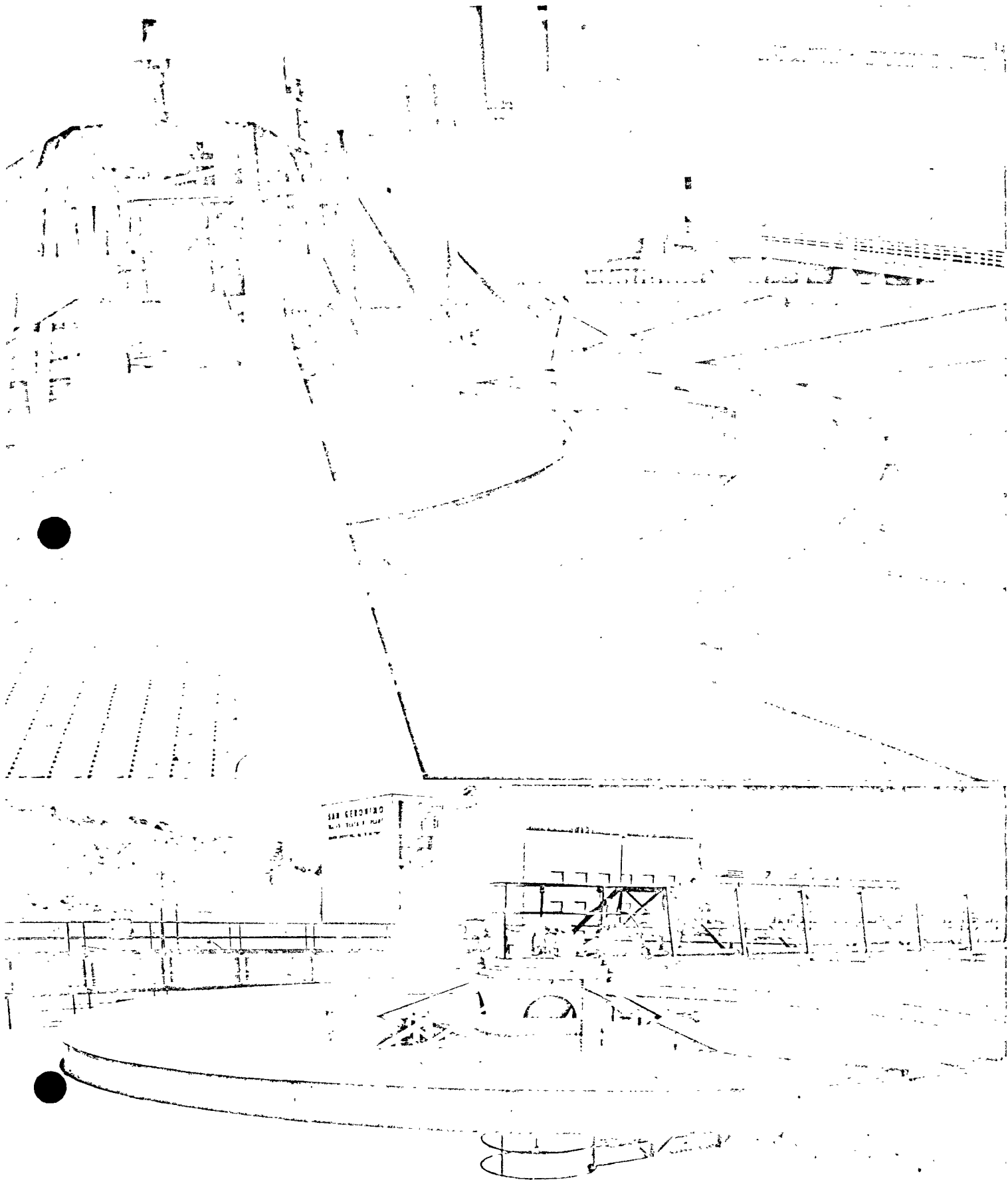
list of users of Eimco Reactor-Clarifier™ treatment units

User	No. Units	Dia. in Feet
Kaiser Steel Corporation	2	90
Fontana, California	1	115
	1	75
	2	55
	1	50
Kaiser Aluminum and Chemical Corporation, Mead, Washington	1	125
U.S. Department of Interior Fairbanks, Alaska	1	32
North Marin County Water Dist. Novato, California	1	55
U.S. Naval Base Guam, Mariana Islands	1	125
Missouri Water Company Independence, Missouri	1	70
PASA Petrochemical Complex Rosario, Argentina	1	85
Ham Tan, Vietnam	1	18
Kontum, Vietnam	1	18
Tuy-Hoa, Vietnam	1	21
Phan-Rang, Vietnam	1	30
Khangh-Hung, Vietnam	1	35
Atlantic Cement Company, Inc. Ravena, New York	1	60
American Oil Company Sugar Creek, Missouri	1	50
Novamont Corporation Neal, West Virginia	1	48
Northern Illinois Water Corp. East Side Plant Champaign, Illinois	4	64
Southern Nevada Power Co. Clark Station Las Vegas, Nevada	1	57
Western Electric Company Millard, Nebraska	1	32
Western Electric Company Lee's Summit, Missouri	1	35
West End Chemical Company Green River, Wyoming	1	32
Albemarle, North Carolina	2	75
U.S. Steel Corporation Columbia-Geneva Division Geneva, Utah	2	135
	2	85
	1	90
Marin Municipal Water District Bon Tempe Plant San Rafael, California	1	125
Goodyear Tire & Rubber Co. Sao Paulo, Brazil	1	35
Union Minere du Haut Katanga Belgian Congo, Africa	1	70
Anaconda Aluminum Co. Columbia Falls, Montana	1	50
Gresik Cement Corporation Surabaia, East Java, Indonesia	1	30
Tata Iron and Steel Co. Ltd. Jamshedpur, India	3	90
H. K. Porter Co. Pascagoula, Mississippi	1	35
Wichita, Kansas	1	50
U.S. Steel Corporation Columbia-Geneva Division Pittsburg, California	2	100
Campbell Soup Co. Sacramento, California	1	115
Northwest Paper Company Cloquet, Minnesota	1	150
Dayton, Ohio	4	120
Anaconda Aluminum Co. Columbia Falls, Montana	1	50
Western Electric Co., Inc. Bell Telephone Laboratories Holmdel, New Jersey	1	30
S.D. Warren Company Muskegon, Michigan	1	110
Jackson County Port Authority Pascagoula, Mississippi	2	110
Independence, Kansas	1	75
General Motors Corp.	1	45
Guide Lamp Division Anderson, Indiana	1	100
Twin City Water Dept. Uhrichsville & Dennison, Ohio	2	60
Vicksburg, Mississippi	2	100
Allied Chemical Corp. General Chemical Div. Painesville, Ohio	1	25
Jones & Laughlin Steel Corp. Cleveland, Ohio	1	70
Union Carbide Corp. Marietta, Ohio	1	30
Republic Steel Corp. Chicago, Illinois	1	80
U. S. Steel Corp. Dravosburg, Pennsylvania	1	150
Caterpillar Tractor Co. Mapleton, Illinois	2	24
Jones & Laughlin Steel Corp. Hennepin, Illinois	2	80
	2	80
U. S. Steel Corp. Gary Sheet & Tin Works Gary, Indiana	1	115
Owens-Illinois Forest Products Div. Orange, Texas	1	115
Union Miniere du Hautkatanga Belgian Congo, Africa	2	70
Alabama Kraft Co. Mahrt, Alabama	1	110
Crucible Steel Co. of America Midland, Pa.	1	48
St. Regis Paper Co. Monticello, Miss.	2	110
Consolidated Aluminum Corp. New Johnsonville, Tennessee	1	82
Western Electric Company, Inc. Indianapolis, Indiana	1	48
Sylvania Electric Company Warren, Pennsylvania	1	40
Continental Can Company Augusta, Georgia	1	160
Kigali, Rwanda, Africa	1	22
New Orleans, Louisiana Algiers Water Purification Plant	1	22
Olin Mathieson Chemical Corp. West Monroe, Louisiana	1	110
Inland Steel Company Hot Strip Tinning Mill Indiana Harbor, Indiana	2	115
Anaconda Aluminum Company Columbia Falls (Conkelley), Mont.	1	50
North American Aviation, Inc. Columbus, Ohio	1	45
Union Carbide Nuclear Company Paducah, Kentucky	1	94

The **Eimco SVG™ Filter** is a simplified, automatic gravity filter with a self-contained backwashing system designed to provide efficient and economical operation for both municipal and industrial plants. The filters can be installed in multiples to meet capacity requirements.

The S V G filter has these major advantages:
. . . Completely automatic in operation.
. . . All piping, pumps and control and regulating valves required for a conventional separate backwash system are eliminated.
. . . No raw water is wasted during the backwash cycle.
. . . Complete control — both hand and fully automatic — is provided.

A single three-way butterfly-type valve provides complete hydraulic control. The valve can be operated either with an electric, hydraulic or pneumatic actuator. While filtering, the valve opens the inlet line and closes the waste line. During backwashing the valve reverses, closes the inlet line and opens the waste line. The backwash cycle is automatically initiated by an adjustable loss-of-head switch in the inlet line (normally set at 5 ft. head loss), or manually by pushbutton. The backwash cycle terminates when a low level probe in the storage compartment is actuated. The backwash rate is adjustable and varies from approximately 24 gpm/sq. ft. to 10 gpm/sq. ft. with an average of 15 gpm/sq. ft. The backwash period lasts for approximately 4½ minutes. The complete control package includes a selector switch which permits manual pushbutton control and control over-ride as well as the normal automatic control. Simplicity of the unit reduces installation costs and space requirements are significantly minimized.

S V G - M
To meet some local and state public health requirements and air space or double wall is needed between filtered and unfiltered water passages and compartments. The SVG-M filter fulfills this requirement by separating the backwash water storage and filtering compartments with an air section and by placing their connecting pipes outside the filtering compartment.
Both types of filters use **Eimco FlexKleen™** distributors. For additional information see page 14.

SVG Filter
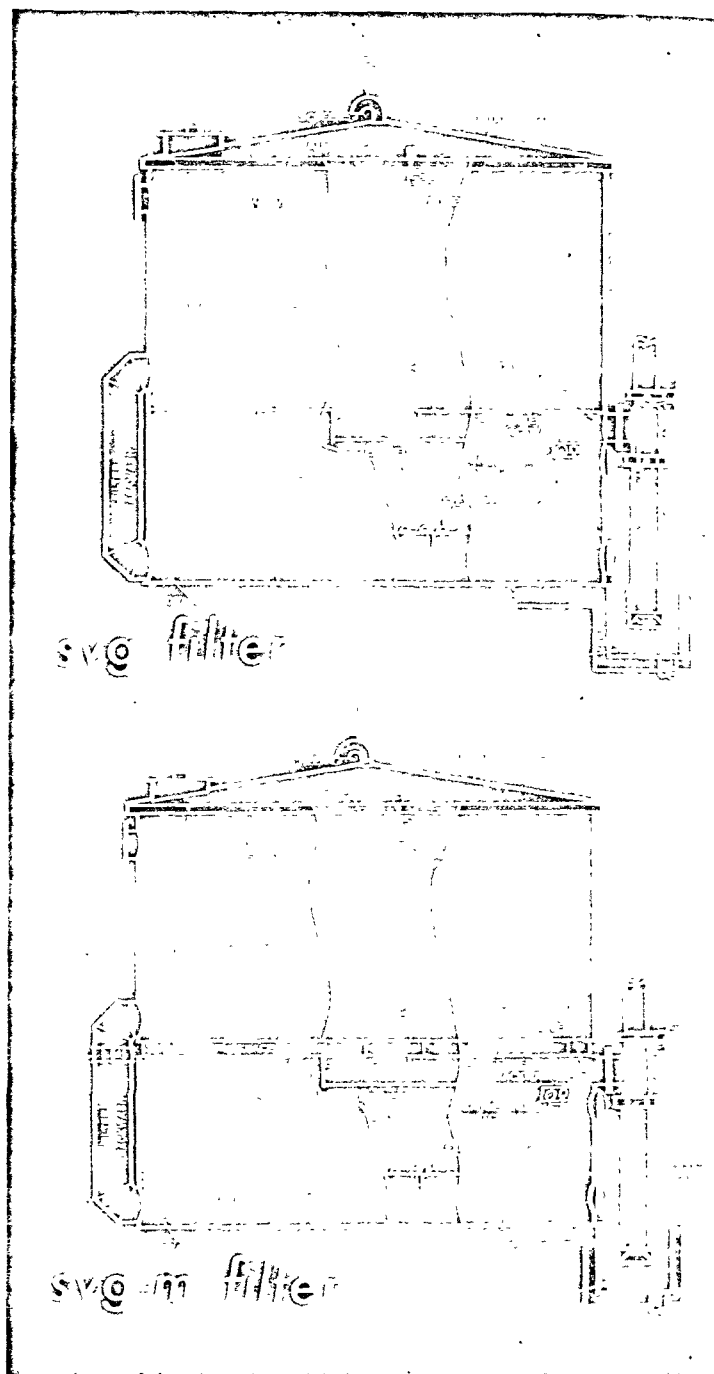
SVG-M Filter

Upper photo
At McGuire Air Force Base near Wrightstown, N. J., the water treatment plant includes 9 S V G-M filters. — *Getter-Green Associates, engineers.*

Lower left
Six Eimco S V G-M filters at the municipal water treatment plant at Galesburg, Illinois. The filters handle 8 mgd removing precipitated iron. — Plant engineering by *Galesburg City Engineering Department.*

Lower right
Two SVG filters polish the water for boiler feed of the Colorado-Ute Electric Association, Inc., steam generating plant at Hayden, Colorado. — *Stanley Engineering Company.*

SVG Filters
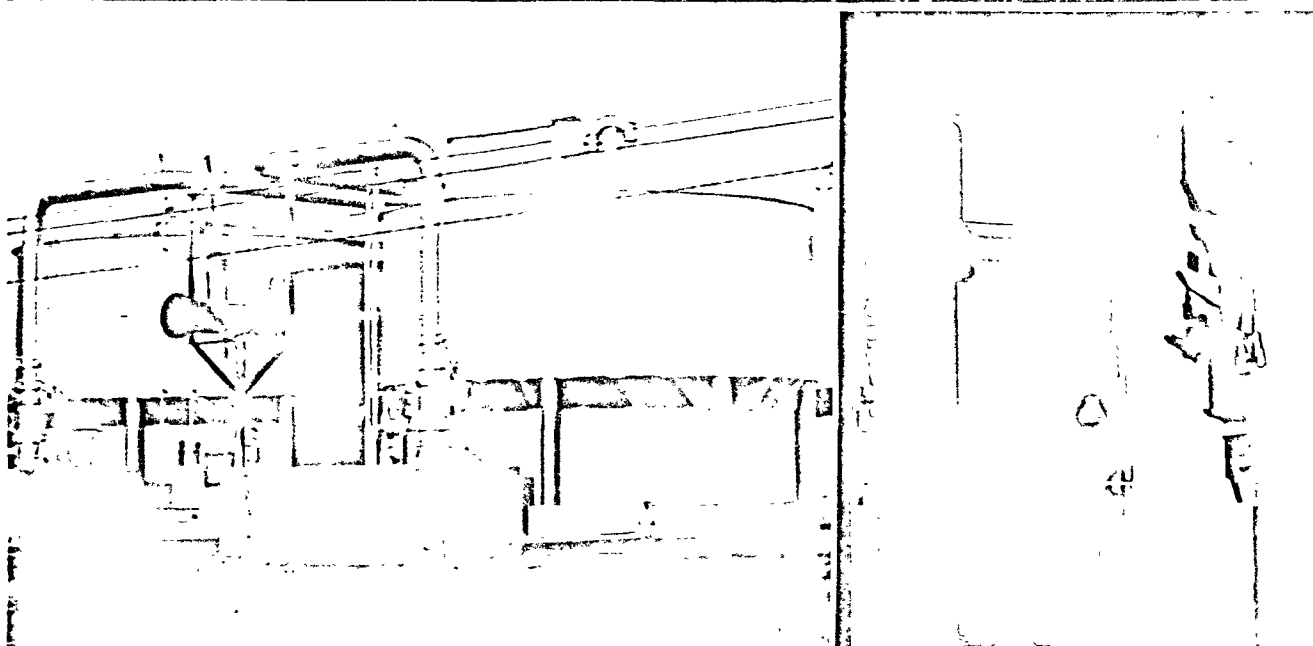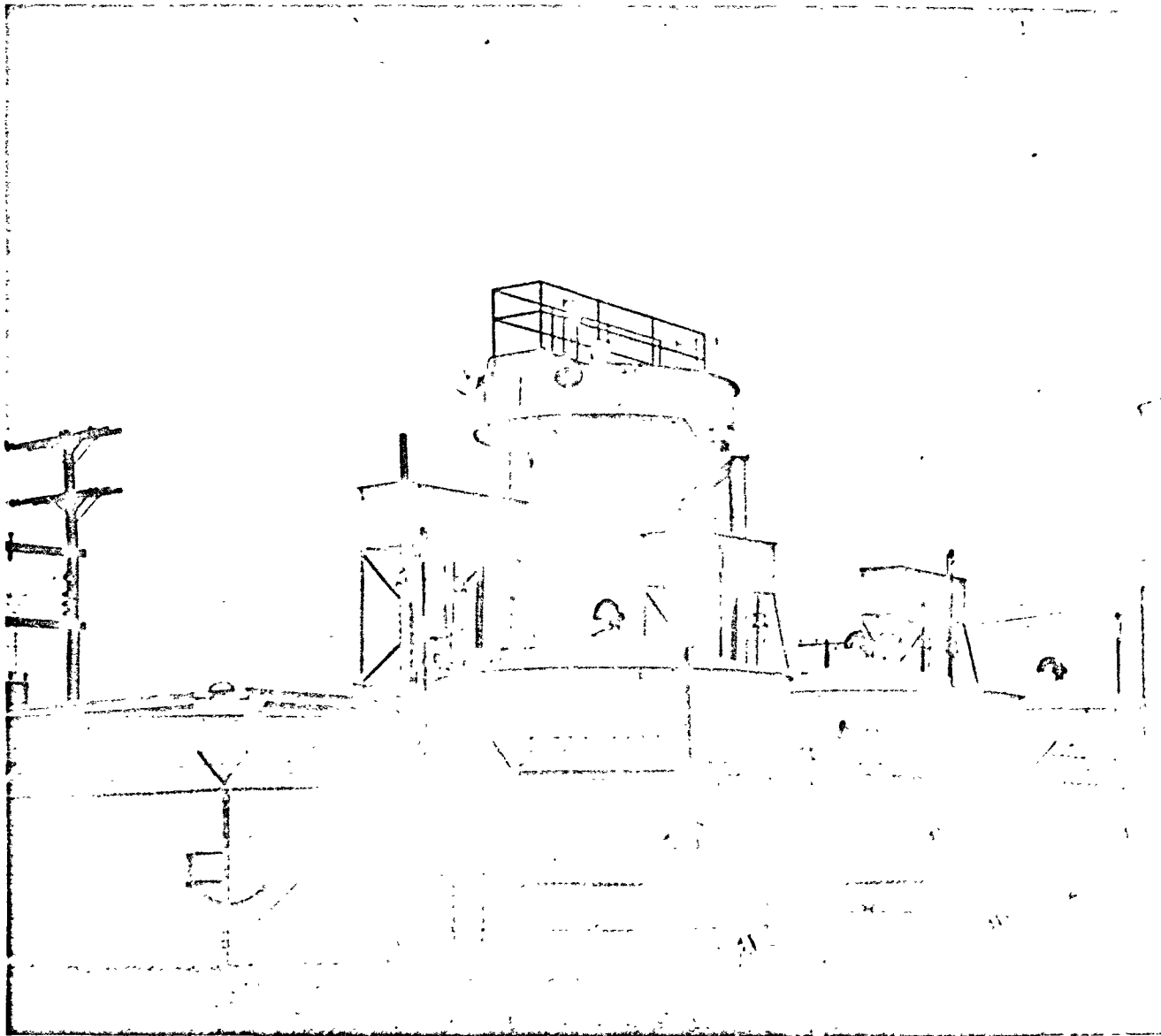
The **Eimco Flocsillator**™ horizontal oscillating flocculating mechanism provides a highly efficient way to gently mix and agitate water for developing floc. All wearing components are above water assuring long life. The flocculating mechanism is particularly suitable for large or small municipal plants where horizontal-flow flocculating is desirable. These mechanisms also can be installed in existing basins to provide flocculation at lower operating costs with a minimum change to existing structures. The paddles of the Flocsillator mechanism travel through approximately 80 per cent of the tank volume and influence 100 per cent of the volume. The action of the lower vertical arms effectively prevents precipitation in the corners of the tank, although there is no high-velocity turbulence to destroy previously formed floc. The Eimco Flocsillator mechanism has many mechanical advantages not found in any other type of flocculator. The drive shaft and bearings are located entirely out of water and are easily accessible for servicing. There are no dry wells, submerged bearings or stuffing boxes to construct or maintain. Mechanisms can be custom designed to fit other basin configurations. Standard Flocsillator mechanisms are available in sizes designed for operation in basins 20 ft. wide by 12 ft. deep and from 16 to 96 ft. or more in length.
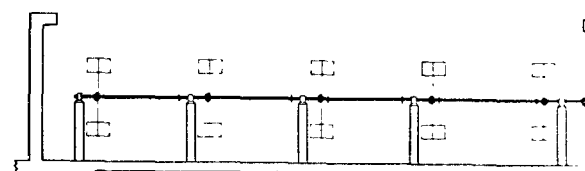
Eimco horizontal shaft (Type HP or Type HT) flocculators are heavy duty units designed for large plants. The variable speed drive unit can be mounted on the roof of or inside a dry well, driving the paddle shaft by means of a roller chain. Two or more flocculators can be operated from one drive unit by sprocket and chain connections in the dry well. Heavy duty construction throughout for years and years of trouble-free service. The HT unit is applicable to high energy application.
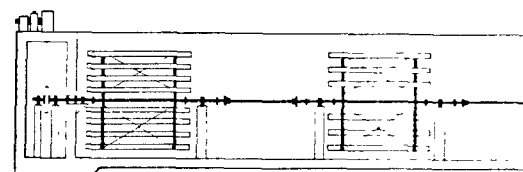
have high circulation capacity using a turbine drive to circulate up to 25 times the initial feed with low power consumption. The blade arrangement is used to vary the flow pattern. Tank sizes up to 38 ft. diameter. No submerged bearings. Applicable to high energy applications.

Vertical paddle flocculators (Type VP) are available in sizes 6 to 50 ft. No dry pits are required. Motor sizes from fractional to 5 hp depending on mechanism diameter and basin depth. Can be installed in series in a single rectangular basin with a surface influent weir and a submerged effluent weir to provide the most efficient flocculation.
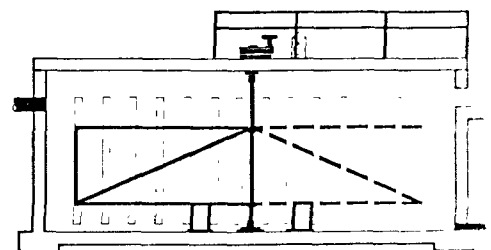
Flash Mixers (Type VT) are available in 1 to 20 hp size vertical turbine design for blending coagulants with raw water and chemical mixing. Also used for neutralization and general purpose mixing.
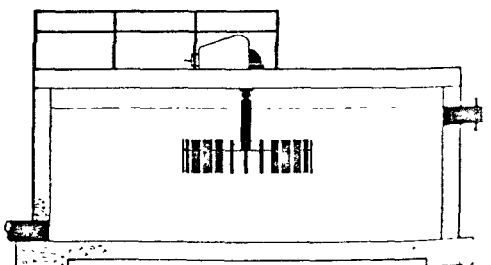
Upper photo
Eimco Flocsillator horizontal oscillating flocculating mechanism at the 400 mgd F. E. Weymouth Memorial Softening and Filtration Plant, Metropolitan Water District of Southern California, LaVerne.
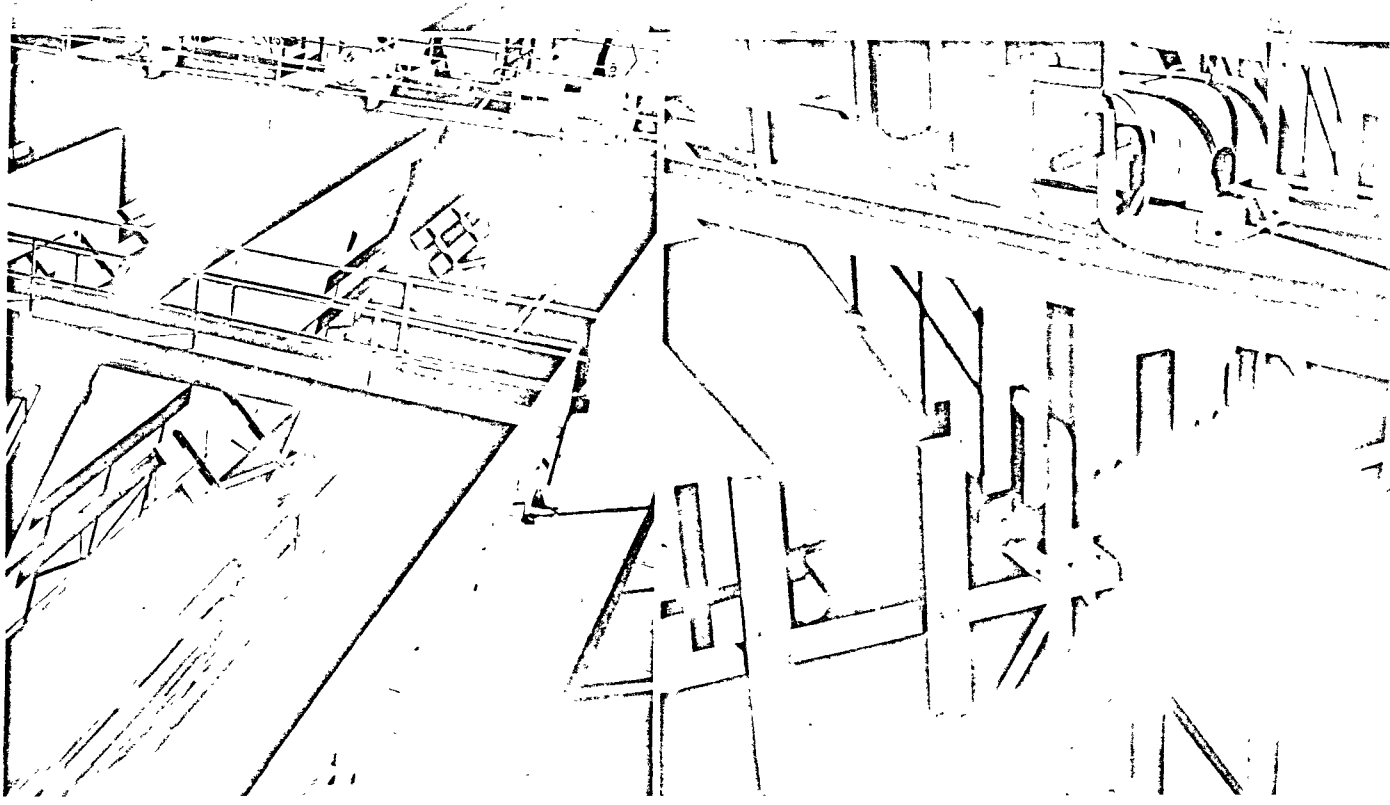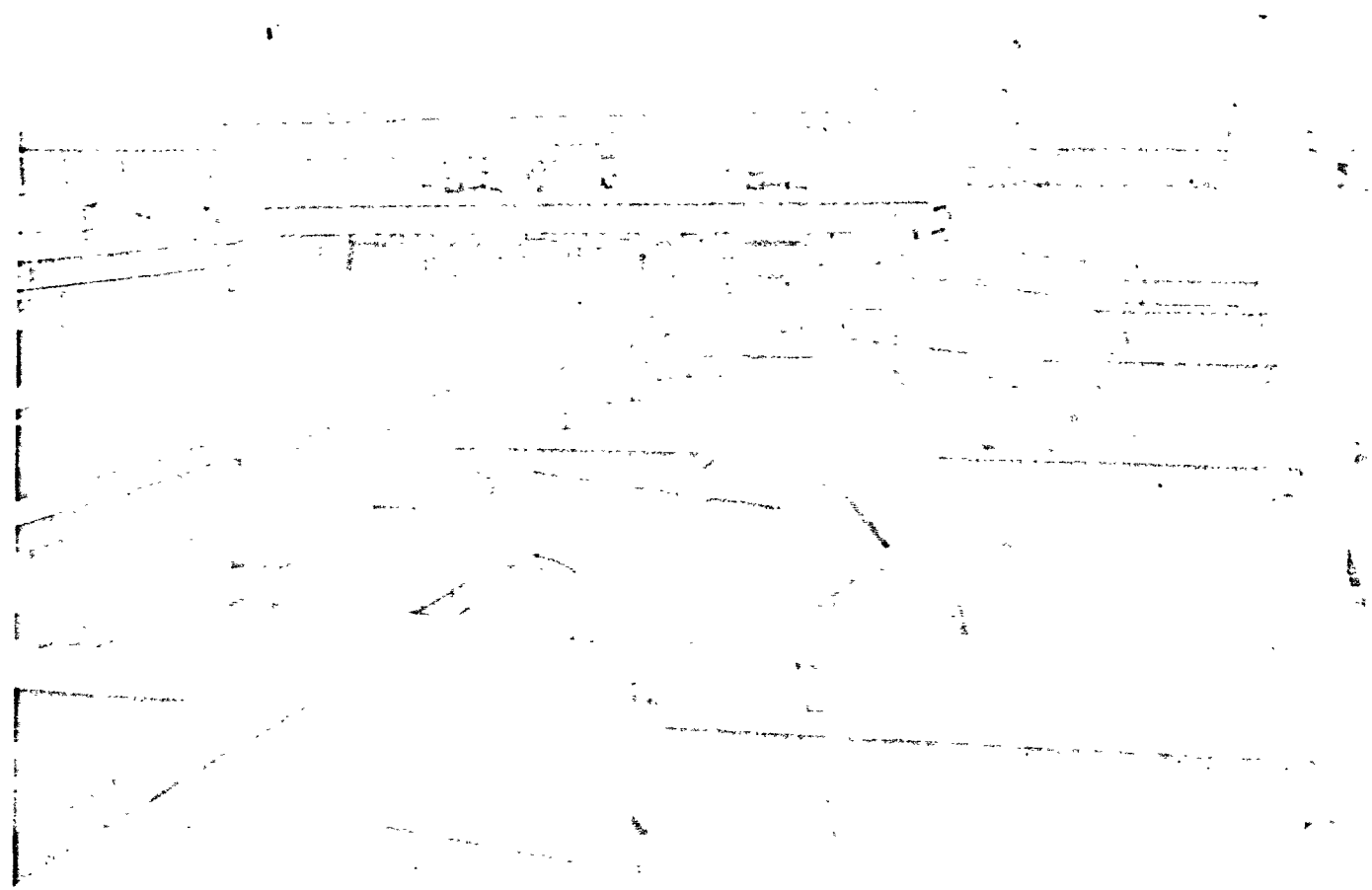
Lower left
One of the sets of horizontal paddle flocculators at the Miramar Water Filtration Plant, San Diego, California. — *James M. Montgomery engineers.*
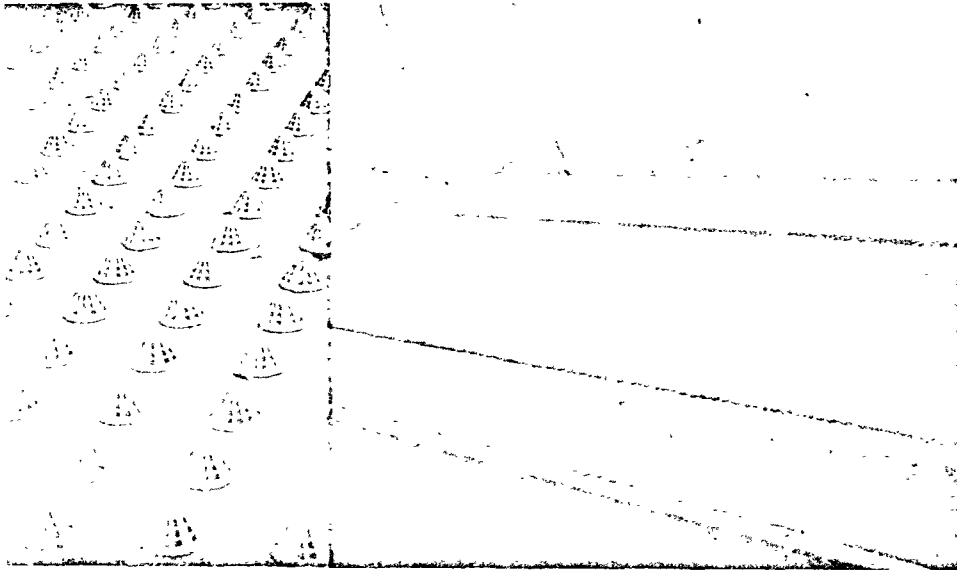
Lower right

Eimco Flocsillator

Type HT

Type HP

Type VP

Type VTR

Flocculators

Air-Water Wash
Dual Media Filters Underdrains

Clarifier:

Dual media double gravity filters which utilize **Eimco FlexKleen**™ nozzles and the **Eimco Air Wash System** have several major advantages for polishing treated water. The FlexKleen nozzles, which are manufactured of non-corrodible plastic with flexible stainless steel screens, are threaded into precast concrete blocks to form a filter bottom. In dual media applications they are covered with a layer of sand and a layer of anthracite. The filter underdrain blocks are 24 in. square, 3 in. thick. They can be supported by circular or square piers on 24 in. centers or on longitudinal beams placed across the filter floor. Threaded plastic inserts cast into the blocks receive the threaded distributors. In the airwash system, the distributors are supplied with plastic tubes which extend below the blocks. The use of air with the backwash provides an especially vigorous wash, with thorough scrubbing of the filter medium. The amount of wash water required is reduced by as much as one half. The agitator eliminates the possibility of "mud balling" tendencies and requires no surface washers or the additional water they use. There is no drifting of gravel or breakthrough caused by surges in backwash. The FlexKleen nozzle is highly resistant to clogging.
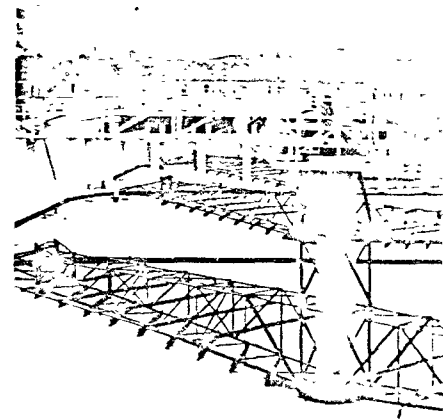
Clarifiers — For pre-sedimentatic of silt and sand or for settling after flocculation. Beam supporte units are available for tanks 20 to 45 ft. diameter; center column and traction types for tanks 30 to 325 ft. diameter. Types availabl for rounded corner square tanks with cross-flow arrangement, sid feed or center siphon feed with conventional overflow. Driveheac are of efficient design with quality construction throughout using high grade materials. Adequate drive gear to machine size ratio equals trouble-free, low maintenance operation.
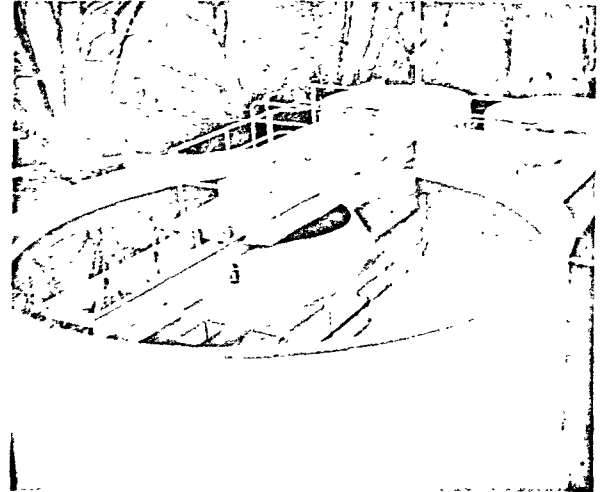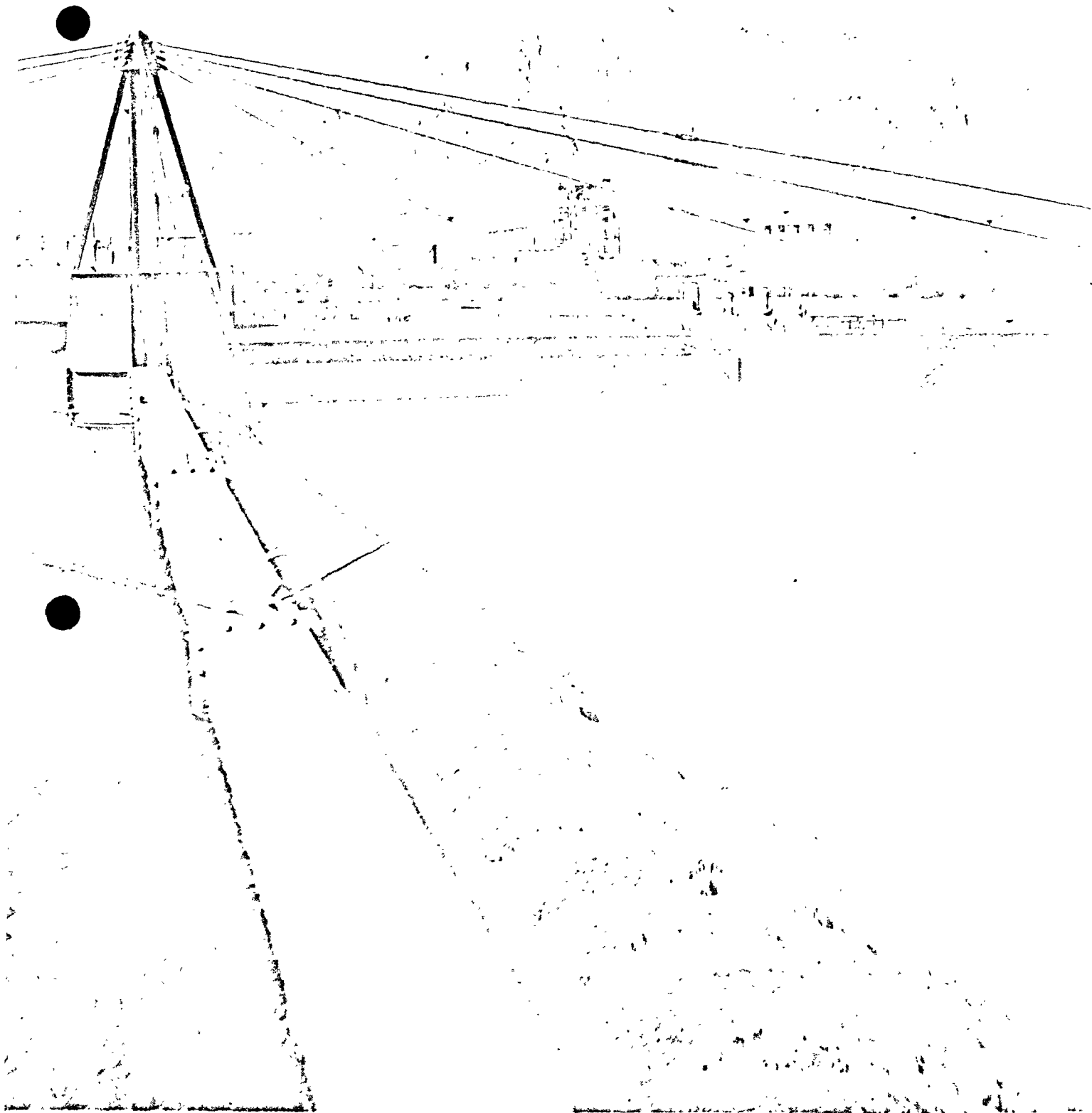
Softening Sludge Dewatering

Package Plants

Eimco Package Plants — Available for gravity operation using S V G automatic sand filters. Eimco offers a complete line of these pre-designed water plants for municipal or industrial use and for boiler feed. Plants are available in sizes from 50 to 1,000 gpm as standard and can be custom designed for larger volumes on request. Many operating efficiencies contribute to lower operating costs. Automatic units are fully dependable.

Eimco offers a package combination to thicken and dewater the sludge which results from softening. Sludges can be dewatered to approximately 65 per cent by weight of dry solids on an **EimcoBelt®** continuous belt filter. The blinding characteristic of the softening sludge presents no filtering problem for an EimcoBelt filter, which operates at all times with a washed, clean medium. Simplified **EdgeTrack™** belt filter operates with minimum of operator attention and provides long belt life. The Eimco thickener used in conjunction with the EimcoBelt filter can be equipped with an automatic raking device to prevent damage to raking arms by the heavy sludge.

The Elm Fork Water Treatment Plant serving the City of Dallas, Texas, has 2 Eimco
Type 2C2 clarifiers, 160-ft., in square basins. — Forrest and Cotton, engineers.

BY: W. H. JOHNSON/THE EIMCO CORPORATION/SAN MATEO, CALIFORNIA

Treatment of Sewage Plant
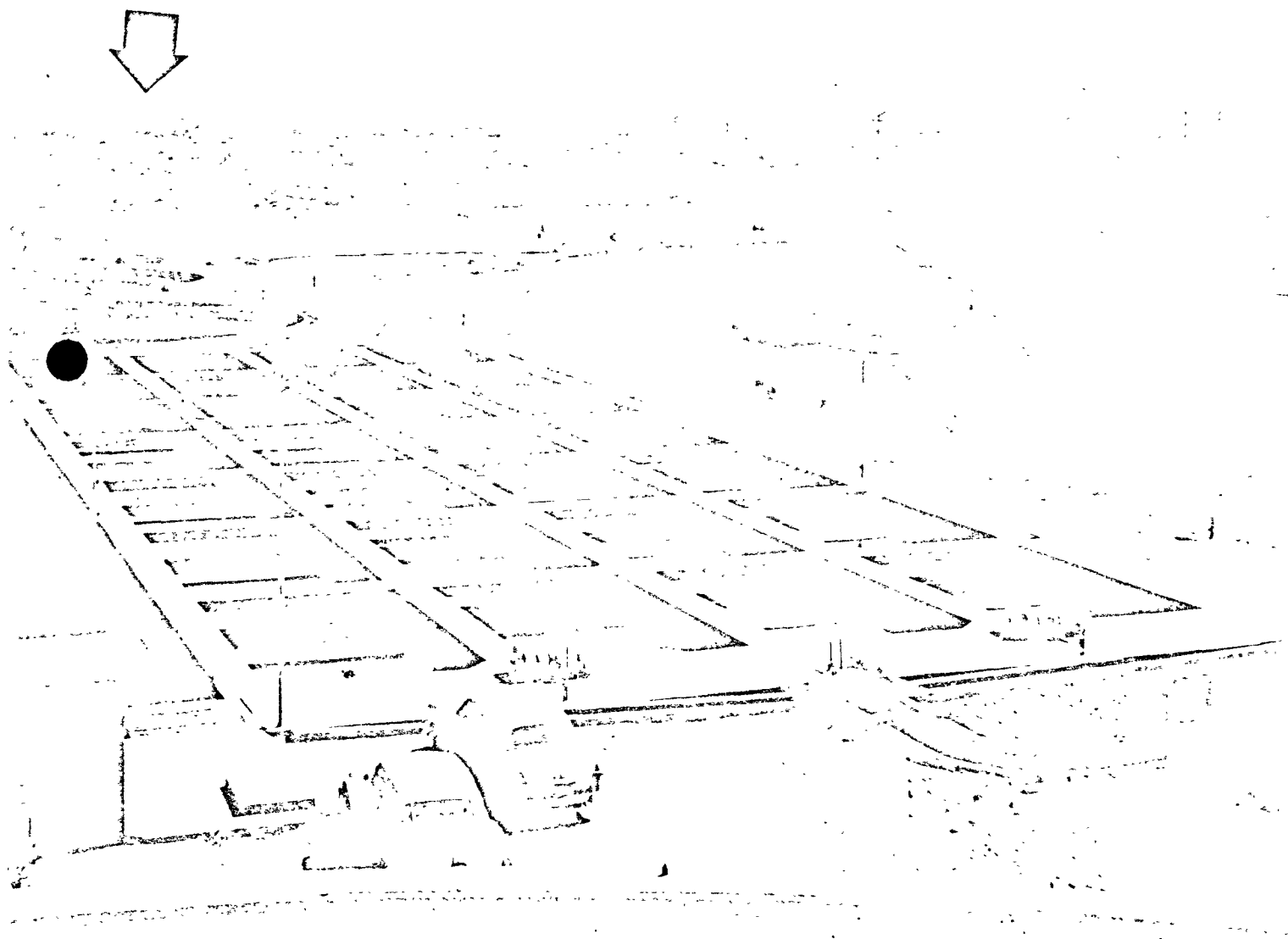# Effluent for Industrial Reuse

Photo No. 1. The Clark County Sewage Treatment Plant near Las Vegas, Nevada. Part of the effluent from this plant is further treated and used for cooling water at the Nevada Power Company's 200 megawatt Clark Station about 1½ miles away (see arrow).

Treatment of Sewage Plant Effluent for Industrial Reuse

BY: W. H. JOHNSON/THE EIMCO CORPORATION/SAN MATEO, CALIFORNIA

Of all the sources of water available to industry, the one most reliable in all seasons and the only one that can be considered as increasing in quantity, and improving in its quality is sewage plant effluent.

The practicability of using this effluent is now firmly established, with over a half dozen industrial plants in the southwest having used it for sufficiently long periods to permit a description of how it should be further treated before it can be safely and efficiently reused.

This water source, when properly treated, should be satisfactory for most cooling purposes and, inasmuch as 75% of all industrially used water is for cooling, this paper will be primarily directed toward discussion of that usage. In particular, it will discuss the special and extra treatment required for the effluent.

The Eimco Corporation is indebted to the personnel at the Las Vegas based Nevada Power Company's Clark and Sunrise Stations and to the Nalco Chemical Company for the generous assistance given in providing much of the data that are used herein. The experience gained during the 3 years of operation of the sewage effluent treatment facilities at the 200 megawatt Clark Station followed by the recent start-up of the newer facilities at the 90 megawatt Sunrise Station, provides much of the data used in this paper. Some comparisons are also included here between other industrial effluent users, particularly those in Amarillo, The Texas Company and the Southwestern Public Service Company, both of whom have provided operating data for this discussion.

Quality is the first consideration in the use of effluent from a domestic sewage treatment plant. Effluents can differ greatly. There are maximum organic content limitations that must be met before it can be called even a partially reclaimed water suitable for further treatment as water.

Biological treatment of the sewage is certainly necessary. Simple primary treatment by gravity settling is not adequate. The effluent BOD (Biochemical Oxygen Demand) should not average over 25 ppm and preferably should be less. Suspended solids should be under 25 ppm and preferably as low as possible. Other constituents, particularly dissolved minerals and synthetic detergents, are not usually influenced by the sewage treatment process and must be taken as they come.

The need for further treatment of the effluent, or reclaimed water as it may be called, will vary with the use to which it is put. It is possible that it may be utilized after only additional settling such as for the once through roll cooling and quenching operations at Bethlehem Steel's Sparrow Point Plant near Baltimore. On the other hand, a very complete treatment may be required such as at Texaco's Amarillo Refinery, where a portion of the reclaimed water is hot lime softened, filtered and further ion exchange softened for use as feed water for low pressure boilers. But since it is recycled cooling water usage that interests us most, it is the required treatment for the protection of the cooling surfaces and towers that we will review in detail.

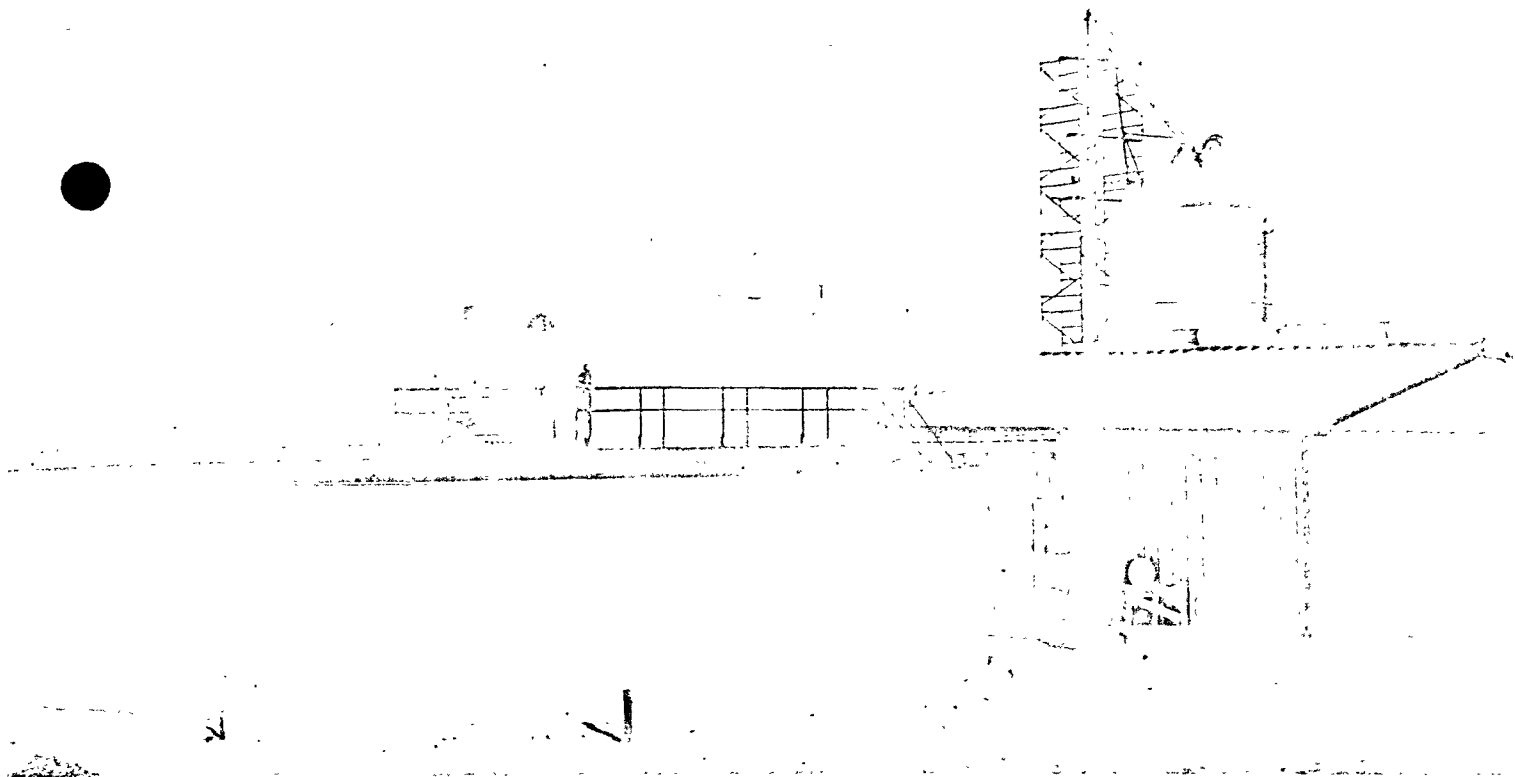It is for the following purposes that a reclaimed water must be further treated so that it may be

3

Photo No. 2. The Nevada Power Company's Clark Station Effluent Treatment Facilities. In the foreground is a 57' diameter cold Lime Treatment Reactor-Clarifier. The chute and hopper over the top of the control building stores unslaked lime.

used with confidence in a recirculating cooling system:

1. Remove orthophosphates.
   Orthophosphates (along with foam causing Alkyl Benzene Sulfonate) come from the synthetic detergents found in increasing quantities in domestic sewage. Only small reductions occur in the conventional sewage treatment processes; therefore, special steps by lime treatment must be taken to eliminate it. Failure to remove this phosphate would result in scaling.

2. Reduce the remaining suspended solids.
   The suspended solids are principally organic and need to be reduced to prevent organic fouling and to keep down chemical dosages. This clarification step and the phosphate removal can be accomplished simultaneously.

3. Kill all bacteria and maintain the residual organic matter sterile. Sterilization is particularly important. Effluent leaving the sewage plant will probably be chlorinated but additional dosages may be needed to keep the water under continuous control.

4. Reduce as much as necessary and practical, hardness, foam, silicates, nutrient matter and other dissolved solids. These are of secondary importance, and should create no great problems but their reduction would be considered beneficial.

Until 1961 Nevada Power Company's Clark Station, used biologically treated sewage plant effluent without further treatment other than chemical additives, in the towers. At that time, the orthophosphates averaged about 17 ppm permitting up to two cycles of concentration. Since 1961 the orthophosphates have increased, at times to over 40 ppm, averaging 35 ppm, making their removal mandatory before the water can be safely used at all. Even so, this sewage plant effluent is equal or superior in overall quality to many raw surface or ground-water sources that are being used for similar purposes in other parts of the country.

The source of the effluent is the Clark County Sewage Treatment Plant several miles away (see Photo No. 1).

In a 1961 enlargement, facilities were constructed at Clark Station (see Photo No. 2) to treat up to 2500 gpm of effluent to remove phosphates and reduce suspended solids. In the spring of 1964 the nearby Sunrise Station was completed. Included in that installation is a 2000 gpm effluent treatment facility (see Photo No. 3). Sunrise Station is connected to both Clark County and the City of Las Vegas (see Photo No. 4) Treatment Plant outlets. Both of the sewage plants use biological filtratilon for secondary treatment as may be seen from the photos. Table 1 gives a typical analysis of the effluent from the City Plant.

CONSTITUENT	SEWAGE PLANT EFFLUENT
Suspended Solids	18 PPM
BOD	20 PPM
Calcium (as $CaCO_3$)	120 PPM
Phosphates (Ortho) as PO_4	35 PPM
Chlorine Residual	0.1 PPM
Temperature (April)	71° F
pH	7.8

Table I. City of Las Vegas Sewage Treatment Plant Typical Effluent Analysis. This reclaimed water is further treated for phosphate removal and suspended solids reduction and used at Nevada Power Company's Sunrise Station.

The treatment at the two Nevada Power Company plants consists of pre-chlorination followed by cold lime softening in a solids contact Reactor-Clarifier using large volume, dense solids sludge recirculation. Removing phosphates and suspended matter by this means has proven to be efficient and economical, rendering the water quite suitable for cooling use and permitting up to five cycles of concentration. Diagram No. 1 illustrates a typical flow sheet of the installations.

Chlorine for sterilization is applied to the sewage plant effluent as it leaves the plant. Additional chlorine is added at the power stations in sufficient quantities to maintain a minimum 1.0 ppm residual. Dosages at the Clark Station presently average 12 to 15 ppm. For a short period each week shock dosages are also applied to the towers.

In the cold lime softening process, phosphates are efficiently removed by adsorption on the precipitated lime sludge. Phosphate content of less than 1 ppm has been consistently maintained in the treated water, and is no longer the controlling factor in concentration cycles. Calcium and total solids concentrations now determine the need for blowdown. Since phosphate removal is by adsorption on the sludge, large volume recirculation of the densest sludge gives the best treatment. In the newer Sunrise Station facilities the Eimco cold lime treatment unit is of the type that carries a low bed or reservoir of sludge which is continuously and internally circulated through the incoming feed. In addition, external sludge blowdown facilities have been piped so that sludge may be either pumped back into the Reactor-Clarifier or excess sludge sent to the drying beds.

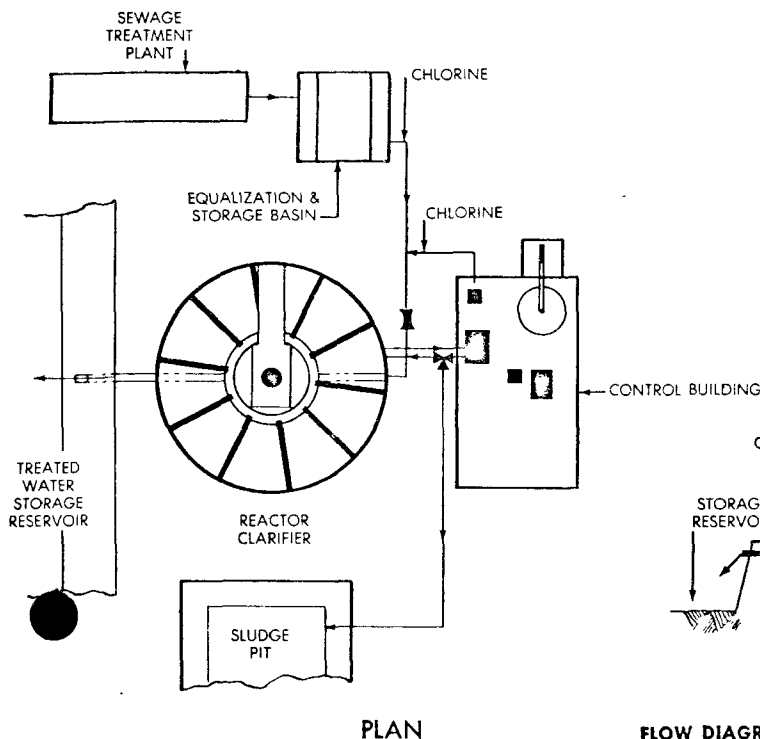Suspended solids or turbidity removal has not been as consistently effective. Organic matter in the effluent does not coagulate easily and sometimes, especially if the solids content is high, good clarity is more difficult to achieve. At clarifier rise rates of 1.25 gpm/sq. ft. satisfactory phosphate removal can be achieved but clarity suffers especially if the sewage effluent contains substantial amounts of suspended matter. Indications are that the best and most consistent clarification results are achieved at rise rates of 0.8 gpm/sq. ft. and a gross detention period of 2½ hours or more.

The older Clark Station Reactor-Clarifier, 57' dia. x 18' deep would normally be rated at 3000 gpm but in order to keep clarity consistently acceptable it treats an average 2000 gpm flow. The newer Sunrise Station unit is 60' dia. x 15' deep and is designed for 2000 gpm.

Failure to maintain a chlorine residual through the Reactor-Clarifier will produce disastrous results, with the sludge becoming septic and completely upsetting clarification.
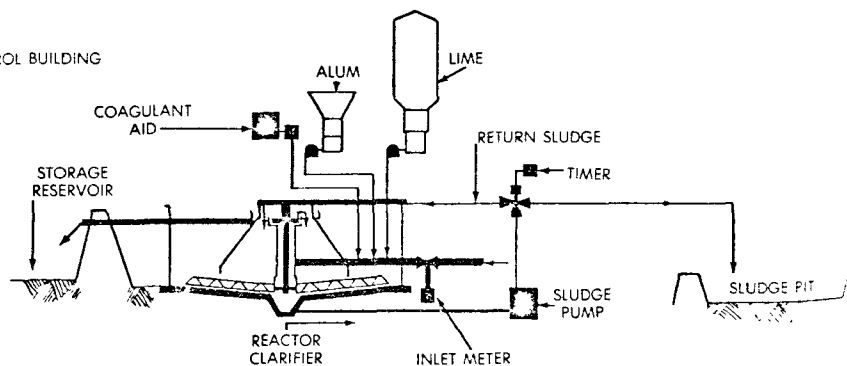
Many types and varieties of coagulants and aids have been tried at Las Vegas but none have yet proven in operation, to give better results than just lime alone. At this writing no other coagulants or aids are used. Most of the time effluent clarities average 5 to 10 ppm, but if incoming suspended solids get high, the overflow may become less clear.

Hardness reduction has been mystifyingly low, while lime demand has been higher than theoretically necessary. Calcium precipitation has been consistently less than that of magnesium. Table II illustrates a typical analysis before and after cold lime treatment. It is generally believed that organics interfering with calcium precipitation account for this phenomena. It is probable that the orthophosphates have some effect too. In tests run by Malina and Tiyaporn (Ref. Fig. No. 1) on the

SEWAGE TREATMENT PLANT

CHLORINE

EQUALIZATION & STORAGE BASIN

CHLORINE

CONTROL BUILDING

TREATED WATER STORAGE RESERVOIR

REACTOR CLARIFIER

SLUDGE PIT

PLAN

FLOW DIAGRAM NO. 1

TYPICAL SEWAGE PLANT EFFLUENT TREATMENT FACILITY TO RECOVER WATER FOR COOLING USE

ALUM   LIME

COAGULANT AID

RETURN SLUDGE

TIMER

STORAGE RESERVOIR

REACTOR CLARIFIER   INLET METER
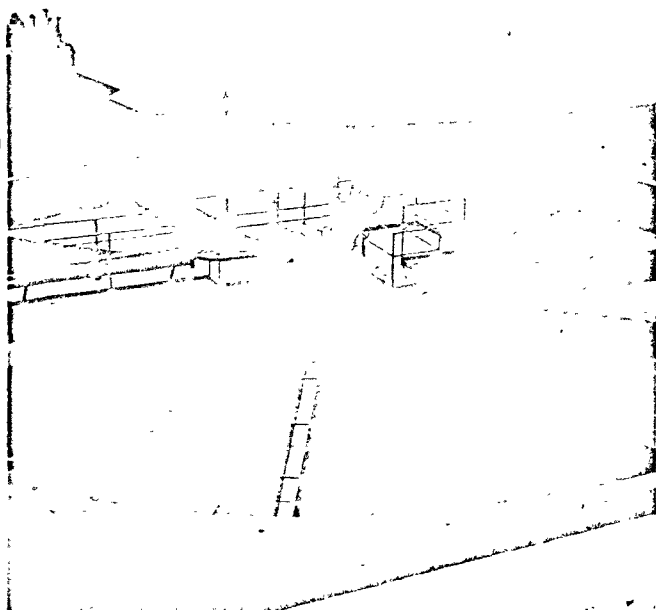
SLUDGE PUMP

SLUDGE PIT

FLOW SHEET

Photo No. 3. The Nevada Power Company's Sunrise Station Effluent Plant includes this 60' diameter solids contact cold lime treatment Reactor-Clarifier with means for large volume, dense solids internal recirculation. Treated water is stored in the basin at the rear.
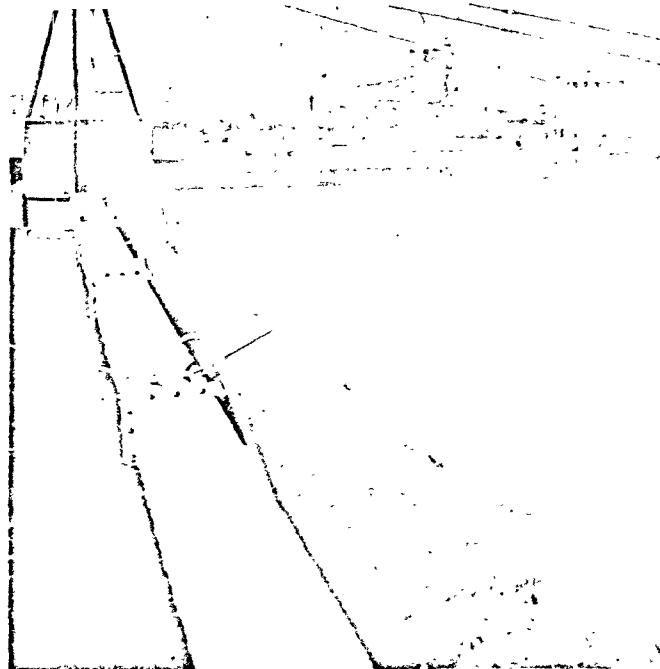

Photo No. 4. Across one of the biological filters at the City of Las Vegas, Nevada Sewage Treatment Plant may be seen Nevada Power Company's 90 megawatt Sunrise Station. A portion of the effluent from this sewage plant is further treated at the power station and used for cooling water.

effect of sodium orthophosphate (and ABS) on hardness removal of a lake water by lime softening, these results were stated — "The residual hardness of the lime treated water increased from 92.9 to 136 mg/1 as $CaCO_3$ as the original phosphate content increased from zero to a concentration of 5.11 mg/1 as PO_4. At concentrations from 5 to 20 mg/1 of Na_2HPO_4 initially added to the water, the hardness of the water after lime treatment was unchanged and remained at about 136 mg/1 as $CaCO_3$". In addition they followed this by stating, "Lime treatment of the water resulted in complete removal of the phosphates, as well as the removal of some of the ABS". Their tests further indicated that the ABS did not interfere with lime treatment of the water.

Constituent	Sewage Plant Effluent To Reactor-Clarifier* PPM	Reactor-Clarifier* Treated Water PPM
Calcium (as $CaCO_3$)	135	115
Magnesium (as $CaCO_3$)	165	105
Total Hardness	300	220
P Alkalinity	240	95
MO Alkalinity	0	165
Sulphates (as Na_2SO_4)	200	200
Chlorides (as NaCl)	200	200
Phosphates (Ortho-PC_4)	14 to 35	Less than 1
Silica	20	10
Total Solids	820	750
pH	7.0	10.0
Chlorine Residual	1	0.3
COD (as O_2)	—	47
BOD	20	—
Suspended Solids	20	5 to 10

Table II. Typical Analysis of Reclaimed Water from Clark County Sewage Plant before and after Cold Lime Treatment. Clark Station — Nevada Power Company — 1962.

* Reactor-Clarifier is the Eimco solids contact cold lime treatment unit

Lime dosages as required to obtain a 2P-M value (two times phenolphthalein alkalinity minus methyl orange alkalinity) of +20 to +40 have given best overall results at Las Vegas. An average dosage of 1.5 lbs. of chemical lime per 1,000 gallons of water treated is required. Treated water pH of 10.0 or slightly above are sent directly to the storage reservoir without acid treatment or carbonation except that naturally obtained through contact with the atmosphere. No scaling has been noted in the Reactor-Clarifier discharge lines or the reservoir. Acid addition of 0.2 lbs. per 1,000 gallons to the circulating tower water maintains the pH between 7.2 and 7.5.

Foaming at Las Vegas has not been a problem since treatment of the effluent began. Formerly, when the effluent was used without further treatment, foam patches were pulled up through the tower fans and blown across the yards. In the upper portion of Photo No. 5 (taken on the day of plant start-up in 1961) may be seen the large white foam areas floating in the water storage reservoir. Within a few days after stable operation was achieved, this foam disappeared (see Photo No. 6) and has not been back since. While no tests were taken to determine the amount of foam causing ABS removed in the cold lime treatment, it does appear that foam is inhibited to a degree. Malina and Tiyaporn's experiments showed a 30 to 60 percent ABS removal by lime treatment of their lake water. In any case foaming should become less of a problem as time goes on since within a few years biodegradable detergents will be in use and they will more likely be removed in the sewage treatment process.

Silica in the sewage effluent runs higher than that of the city water supplies, but its reduction by 20% to 60% in the cold lime treatment takes it down below its former level. The relatively high
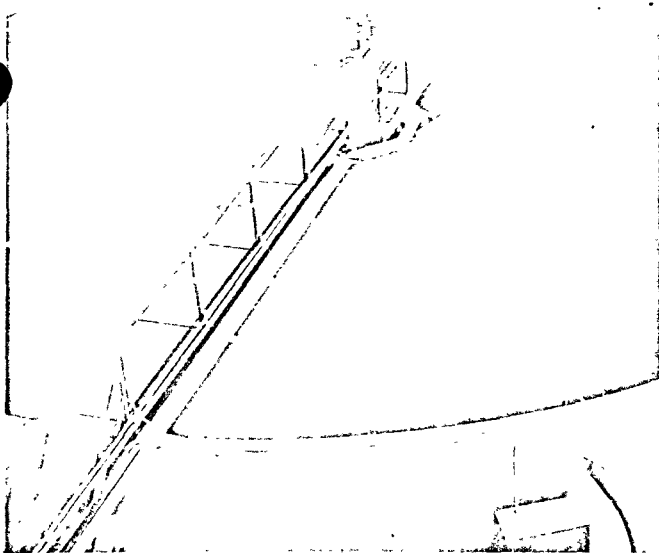
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Photo No. 5. Before. Photo taken at start-up time of the Clark Station effluent treatment facilities in 1961. Note the white foam still remaining in the storage basin in the background. After a few days of operation, the foam disappeared and has never returned. (See Photo No. 6) The unit shown is a 57' diameter solids contact cold lime treatment Reactor-Clarifier.
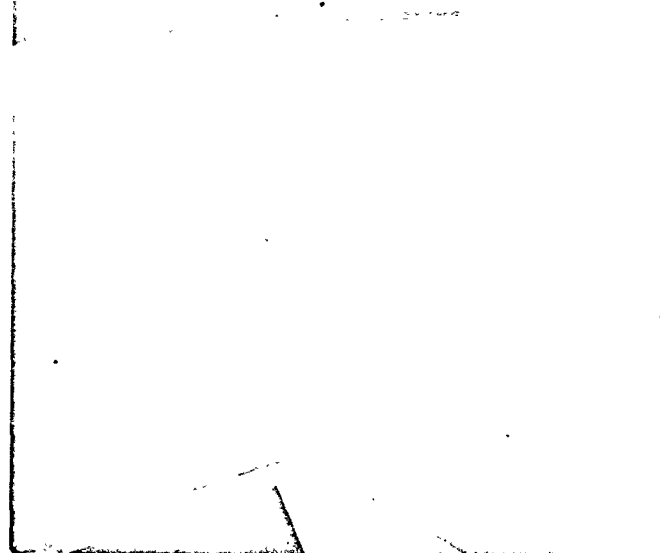

Photo No. 6. After. Clark Station storage basins in 1962 after effluent treatment facilities had been in operation for some time. Note the complete absence of foam, slime or algae growth.

magnesium hydroxide precipitation no doubt accounts for this.

Overall operation and results have been quite satisfactory. The treated water is stored in an open reservoir without developing slime or algae growths. Copper sulphate treatment has not been found necessary as long as chlorine residuals are maintained through the clarifier. Tube inspection after 3 years has shown only slight scaling and it has not been organic. The average number of cycles of concentration has been five, with the limit determined by calcium and total dissolved solids concentration. Only the usual tower water control chemicals have been required — acid (for pH), polyphosphates, algicides and periodic shock chlorination.

Experiences with reclaimed water at Amarillo have been generally similar to those at Las Vegas but with some variances. The main differences are these:

Amarillo's Water Reclamation Plant uses the activated sludge process of sewage treatment followed by 3 days storage at the plant. This storage serves to further oxidize, clarify and equalize the effluent before it is sent to the two industrial users — Southwestern Public Service Company's Nichols Station and the Texaco Refinery. Table III gives a comparative analysis of Amarillo's City Water, raw sewage, reclaimed water and treated effluent.

Chlorine is applied in sufficiently heavy doses at the sewage plant to maintain 1 to 5 ppm residual in the effluent at the time it reaches the user. An average of 9 ppm is used to do this.

Clarity results are similar to those at Las Vegas, however alum has been found to be beneficial when used at the rate of 30 ppm. During difficult periods small amounts of Separan NP10 have also been of

some benefit. But there are times when good coagulation can not be achieved and clarity becomes poorer. These occasional bad periods are apparently not of great overall significance.

At the Public Service Plant, pH is reduced immediately after treatment to 9.0 by sulphuric acid, to prevent scaling in the lines and storage tanks.

A possible reason for a higher scaling tendency at Amarillo than at Las Vegas may be because they treat to a higher phenolphtalein alkalinity, aiming for a 2P-M value of between +50 and +90. Also there are not as many provisions for recirculation of dense bottom sludges as at Las Vegas, which aids in stabilizing the water.

Constituent	Amarillo City Water	Untreated Raw Sewage	Sewage Plant Effluent	Cold Lime Treated Water
Calcium ($CaCO_3$)	39 PPM	110 PPM	110 PPM	100 PPM
Magnesium ($CaCO_3$)	37 PPM	140 PPM	140 PPM	80 PPM
Sodium	27 PPM	—	110 PPM	—
Iron	0	—	0.3 PPM	—
M. Alkalinity	225 PPM	367 PPM	334 PPM	270 PPM
Hardness ($CaCO_3$)	244 PPM	250 PPM	250 PPM	180 PPM
Silicate	56 PPM	—	79 PPM	45 PPM
Ammonia Nitrogen	0	25 PPM	20 PPM	—
Nitrate Nitrogen	1 PPM	0	2.30	—
Phosphate (PO_4)	0	35 PPM	27 PPM	1.0
Chloride	14 PPM	162 PPM	83 PPM	—
Sulfate	36 PPM	—	78 PPM	—
Total Dissolved Solids	356 PPM	671 PPM	557 PPM	—
Suspended Solids	0	236 PPM	11 PPM	5 PPM
Biochemical Oxy. Demand	0	275 PPM	10 PPM	—
Chlorine Residual	0.2	0	5.0	1.0
pH	7.5	7.5	7.7	10.3

Table III. City of Amarillo, Texas Comparative Typical Analysis of City Water, Raw Sewage and Treated Effluent from the City's Water Reclamation Plant. This Reclaimed Water is further treated and used by the Texaco Refinery and Southwestern Public Service Company's Nichols Station to give the typical results shown in the last column.

Photo No. 7. Clark Station cooling towers using treated sewage plant effluent average five cycles of concentration. No foaming problems are experienced.

Two special problems involved in the use of sewage plant effluent are (1) its residual organic content and (2) its relatively inconsistent character. Raw sewage is quite variable in both its quality and quantity. It varies by the hour, by the day and even by the month. A secondary sewage treatment process will tend to even out some of the fluctuations with BOD and suspended solids removal ranging from 75% to 90% throughout any day, varying both with flow rate and the raw sewage quality. It will not appreciably affect mineral content, hardness, alkalinity, temperature or pH. While it is reasonable to expect to average less than the maximum permissible BOD and suspended solids in the effluent, there will be those fluctuations along with the other quality variations that may affect the effluent user's additional treatment steps. This is one of the important points that must be taken into account when reclaimed water is being considered.

To overcome these variations, in part at least, an equalization and storage basin should be provided if possible. Its function would be to even out the effluent quality and provide stand-by storage as well. A basin with a day's capacity or more, with proper inlet and outlet provisions to prevent short circuiting would help considerably. The benefits of further settling and oxidation will also improve quality. By including a means of aerating and mixing a portion of the storage tank contents additional organic content reductions could also be obtained.

At Las Vegas there was no such basin for the County plant. However an approximately one-day capacity pond is now nearing completion. At the City's plant a small basin has been provided. At Amarillo's City Water Reclamation Plant, two lined basins with a total of 3 day's storage capacity are part of the water reclamation facilities. In addition to the equalization and storage benefits, the Amarillo basins account for a 5 to 30 ppm greater BOD reduction and a 5 to 15 ppm increased suspended solids reduction.

At an industrial plant in Mexico where these basins were not wanted and yet a consistently high quality flow was desired, gravity sand filtration was provided prior to cold lime treatment. This too, is an effective means.

Without such filters or basins, the effluent user's treatment plant operators must be prepared to exercise more careful and regular observation and control of the water treating systems.

In summary then, we can say that in over 3 years of experience at its two stations, the Nevada Power Company has found that sewage plant effluent produces a water entirely suitable for industrial cooling use when further treated by continuous chlorination and cold lime in a solids contact treatment unit.

The sewage must be biologically treated to as great a degree as practical and if possible the effluent should be equalized so as to even out the continuously variable qualities of the effluent.

Excellent phosphate removals and reasonable clarities can be expected but relatively little in hardness reduction. No other serious problems or foaming have come about from dissolved matter. The treated water when adequately chlorinated may even be stored in open reservoirs with no particular slime or algae problems.

In view of the increasing demand for water by industry and use of over 75% of its water consumption for cooling, it is significant that sewage plant effluent as an ever increasing source of supply is proper for this use.

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    Joseph F. Malina Jr., and Supote Tiyaporn
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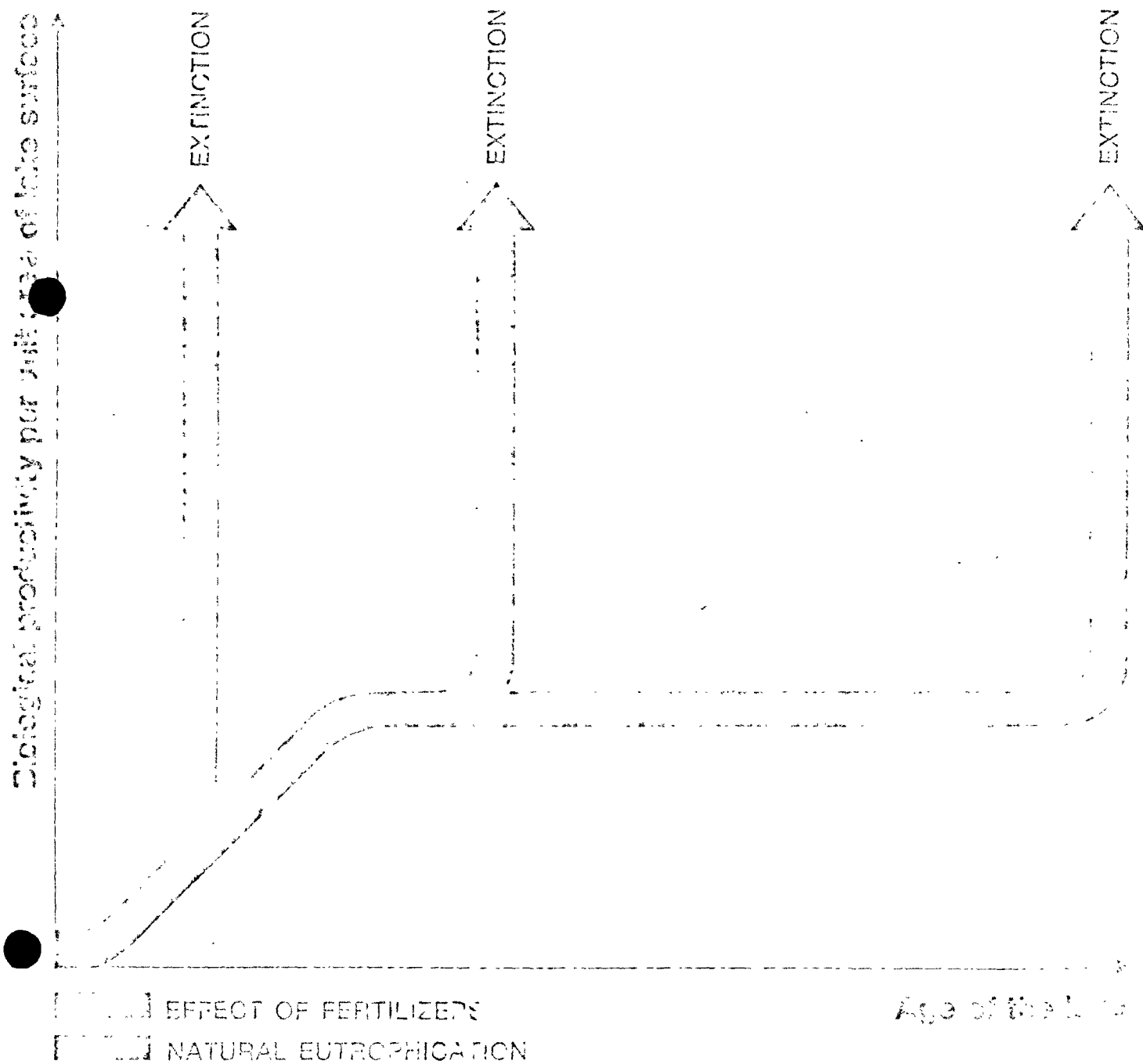
Introduction

In its infancy a lake may have been a barren body of water lacking the critical nutrients that support aquatic life. With the passing of time, wind and rain transported these necessary nutrients and the lake took on life. After an initial surge of productivity the lake entered a long period of dynamic equilibrium. It is during the plateau in productivity that the lake offers the most benefits —commercially and aesthetically.

The natural useful life of a lake should be measured in tens of thousands of years, however the contamination of man is causing extreme premature extinction of many waters. For the last half century man has observed the changes brought about by over fertilization. Passive realization has now turned to active concern for effective methods to control this accelerated aging. A realistic solution is the Dorr-Oliver Phosphate Extraction Process.

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hosphate Extraction Process

Orris E. Albertson, Marketing Manager
nd Robert J. Sherwood, Marketing Engineer
ater Management Systems, Dorr-Oliver Inc.

e sanitary engineer has all the tools necessary to
ectively eliminate the gross quantity of organic
lutants—that is the initial 90-95% of the BOD and
spended solids. Often, however, receiving waters
I require effluent beneficiation. There are several
ms this beneficiation may take — increased solids
noval, higher BOD and COD removal, and phos-
ate and nitrogen reduction.

One of the fundamental problems in stream pollu-
n is the accurate prediction of the effects of treated
mestic and industrial waste discharges upon re-
ving waters. Heretofore, the greatest emphasis
s been assigned to the influence of the organic
ste residuals upon the oxygen resources of the
eam. However, the oxygen yardstick, established
the 5-day BOD concept, although frequently use-
cannot satisfactorily account for the total pollu-
nal potential of a treated effluent.

Why remove phosphorus? It is only about 6 mg/l
final effluent. One cannot see, taste or smell it in
nal effluent. Perhaps it is these very characteris-
. that have slowed the efforts of many researchers
oring the fertilization into its proper perspective.
The magnitude of the pollution problem associated
n the phosphorus in final effluent is not generally
mprehended. For example, a final effluent con-
ning only 10-15 mg/l BOD and TSS and 6 mg/l
phosphorus has a growth potential as follows:

$$\text{mg/l P } \frac{\text{Light, nutrients}}{\text{carbon}} \quad 400\text{-}600 \text{ mg/l algae}$$

$$00\text{-}600 \text{ mg/l TSS } \cong 400\text{-}600 \text{ mg/l COD}$$

hus, the phosphorus can produce a growth re-
nse that may equal the organic pollution load in
raw sewage This realization has spurred gov-
nental efforts to develop systems which would
vide the sanitary engineer with means to remove
phosphate from plant effluents

here has been an increasing concern over the
of economical means of abating excessive or-
c and nutrient enrichment of receiving waters
ch then develop prolific growths of algae. These
e growths have aesthetically deteriorated the
ace waters and often limited its economic value.

Algae scums discolor the water and, upon decom-
position, release foul odors to the water and to the
atmosphere. Filter clogging by algae cause diffi-
culties in the purification of water and increase water
treatment costs. In addition to the productivity re-
sponse to fertilization, severe secondary pollution
results from the additional oxygen requirements
associated with the ultimate decomposition of the
algal organic material.

Sanitary engineers have long recognized that do-
mestic sewage and some industrial wastes are a
rich source of the critical nutrients which cause algae
to flourish. The degree of eutrophication, and hence
the severity of subsequent water quality problems,
is largely dependent on the supply of inorganic nitro-
gen and phosphorus. Chemical control has been
employed to prevent excessive algae blooms. Such
treatment included the periodic application of copper
sulphate or other algicides, or diversion of nutrient
rich wastes to less sensitive or less valuable receiv-
ing waters, or a combination of these measures.

These control procedures have obvious limitations
and will not be broadly utilized. Cost and subsequent
toxic effects of the best available algicides precludes
their use for continuous control of most surface
waters Furthermore, the effort of most algicides is
only temporary and does not attack the real cause
of the problem Recently, the interest in developing
a method of waste treatment which would remove
offending nutrient elements before effluent discharge
has been renewed.

Sawyer[1] established that phosphorus removal
offers a practical and effective way of controlling
algae growths in most surface waters. Other investi-
gators,[2][3] particularly in Europe, have investigated
the role that nitrogen, in various forms, plays in the
growth of algae Much of the earlier efforts in Europe
have been directed towards removing nitrogen from
the receiving waters Only recently have their investi-
gations turned to the phosphates and the role that
phosphate and nitrogen together have on the growth
rate and quantity of algae.

Phosphorus removal may be accomplished by bio-
logical or by chemical means. Both approaches are
directed toward converting soluble and colloidal

phosphorus into recoverable insoluble material. Of the two, chemical coagulation has received the greatest attention and several effective, but costly, chemical treatment methods[4] [5] [6] [7] have been proposed. To date, most of these chemical treatment methods have investigated removal of inorganic phosphate from sewage treatment plant effluents In other investigations, the use of iron and aluminum salts has been evaluated[8] when added directly into the aeration tank of the activated sludge system. Interest has been recently reviewed in the application of biological mechanisms for achieving higher phosphate removals without adding a tertiary stage.

The mechanism of phosphate removal by chemical coagulation is not well understood Theoretically, phosphorus may be removed from solution through precipitation as an insoluble salt or by absorption upon some insoluble solid phase. Available experimental evidence indicates that both mechanisms may be operative, particularly at low residual phosphorus concentrations In the case of lime coagulation. it appears that the principal mechanism is that of precipitation as insoluble calcium phosphate salts With iron salts and alum, absorption upon hydrated oxide floc-particles appears to play a major role along with the formation of an insoluble salt Pilot plant data indicate that poor floc settling properties may require much lower clarifier overflow rates than commonly employed in sewage treatment[9].

Figure 1, developed from the reviews of Nesbitt[10] and Clesceri[11], summarizes phosphate removal by a number of investigators. This figure indicates that cost appears to be the major limitation to the application of present chemical coagulation processes. Based on available information, the cost of chemicals alone would range from $20 to $80 per million gallons

FIG.1—REPORTED COSTS FOR PHOSPHATE REMOVAL

O < 4.0 mg/l p
△ = 4.1—6.0 mg/l p
□ = 6.1—8.0 mg/l p
● > 8.1 mg/l p

Chemical Dosage—mg/l
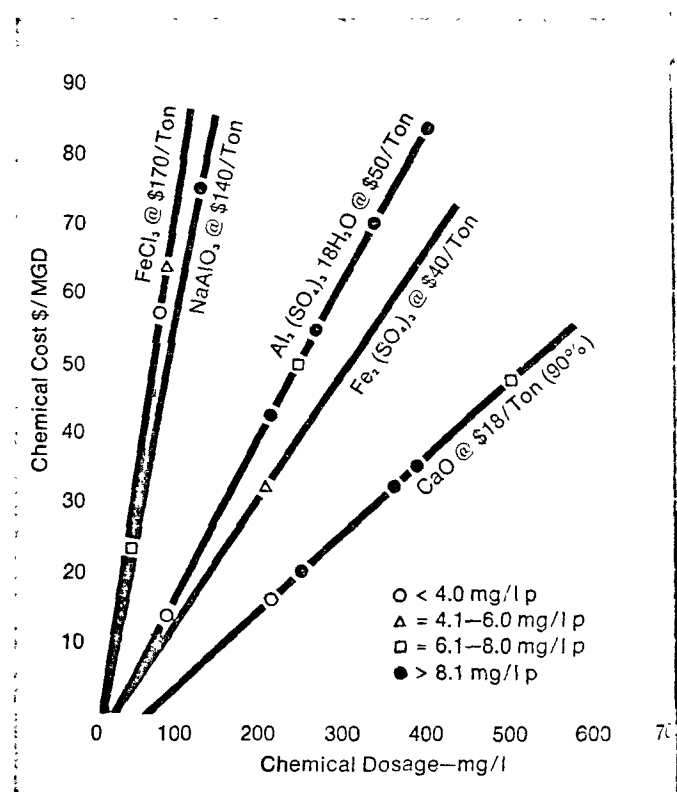
of sewage treated for an 80-95% reduction in phosphorus.

In addition to the operating cost for chemicals, there are additional structures required for tertiary treatment. Cost of operating personnel, maintenance, and disposal of the chemical precipitated sludge can greatly inflate costs shown in Figure 1. Lime treatment costs look attractive, but the cost of neutralization is not included and it could increase the cost $10-$15/MG.

Under suitable conditions, soluble phosphate may combine with a number of substances ordinarily present in sewage or with added compounds to form relatively insoluble complexes. The basic relationship of residual phosphate concentrations as a function of lime dosage normally is similar to a monomolecular curve. Precipitation with lime will generally follow a curve, as shown in Figure 2, when the bulk of the phosphates in the waste is orthophosphate. The actual chemical dosage required to meet a specific residual phosphate will be dependent upon the calcium hardness and the total alkalinity of the water. However, depending upon the clarification characteristics of the calcium phosphate suspension, the actual removals could be much poorer.

wyer[12], in his discussion of lime treatment of raw waste, indicated that the lime requirement, to reach a pH of 11, is about 2.0 to 2.5 times the alkalinity of the waste. He also stated that there will be a variation in the characteristics of phosphate removal from raw sewage vs. phosphate removal in the final effluent after activated sludge treatment. In the raw waste, the orthophosphate may constitute only 40-60% of the total phosphate, but it will usually constitute over 90% of the total phosphate in the effluent. In the influent raw waste, a significant portion of the phosphate will be organic phosphates and polyphosphates, which are more difficult to remove with lime than the orthophosphates.

Other agents may be used to produce insoluble phosphate compounds, but these may not follow the removal characteristics produced by lime addition. For example, the aluminum and iron compounds react to form metallic phosphates, $AlPO_4$ or $FePO_4$, which are insoluble under specific pH conditions. However, the use of aluminum or iron ions for phosphate removal as a tertiary treatment step produces vast amounts of sludge. From the standpoint of chemical cost and the sludge handling problem, the use of these compounds has not been considered economical. Alum, by itself, can produce as much as 10,000 gal/day of additional waste sludge per M( )of waste water.

FIG. 2—REMOVAL OF PHOSPHATES WITH LIME
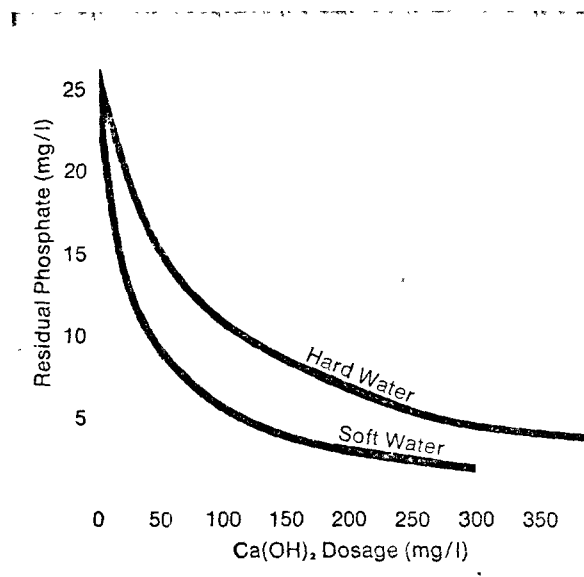
More recently, investigators[8][13][14] have added Fe^{++}, Fe^{+++}, and Al^{+++} ions directly to the activated sludge culture. In this application, there is a semi-stoichiometric reaction. That is, the metallic ion dosage is a function of the phosphorus concentration. Strict stoichiometric relationship is not maintained and the dosage is reported to be about 1.1 to 2.0 times the phosphorus requirement. On this basis, the chemical cost could be quite practical in those wastes having a low phosphorus content.

Clesceri[11], in his summary on phosphate removal, reported that the mixed liquor suspended solids also has an effect on the chemical dosage Normally, the higher the mixed liquor suspended solids concentration, the higher the aluminum or iron dosage that is required for a desired phosphate removal. With good clarification, the addition of metallic ions to the aeration basin can be expected to reduce the phosphorus level to less than 0.5 mg/l in the final effluent. It is not possible, however, to utilize anaerobic digestion for handling sludge containing phosphate precipitated by ferric ions because the anaerobic system will reduce the ferric to ferrous and the phosphate will again become soluble.

3

Biological removal of phosphorus

The concept of removing nutrients biologically cannot be considered as new or unique In the activated sludge system, oxygen is supplied for the use of micro-organisms to convert carbon, hydrogen, sulfur and nitrogen from their reduced level to a higher oxidized state The oxidation of these basic elements is carried out by micro-organisms that retain energy from these reactions for the synthesis of new cellular protoplasm In any actively growing system, nutrient materials are continually extracted from the environment and converted to cell tissue The rate of nutrient removal is proportionate to the rate of net cell tissue synthesis, and the cell tissue composition of nitrogen and phosphorus Bacterial growth rates vary greatly with the type of organism and with the species, but the mixed microbial culture provided by the activated sludge process appears to be most effective biological system based on observed rates of phosphate removal. Also, it is doubtful whether a select culture, having a high phosphorus requirement, could be maintained pure in a waste treatment system.

The nutrient removal efficiency of present-day activated sludge systems is dependent upon the carbon-to-phosphorus ratio of the waste being treated. Thus, the treatment of waste with high ratios of assimilable carbon to phosphorus — for example, sugar wastes generally result in high efficiencies in nutrient removal or may, in fact, require nutrient addition The use of phosphate in detergents, and their expanded applications, has greatly reduced the carbon-to-phosphorus ratio in domestic wastes In turn, the phosphorus elimination by the activated sludge culture suffered correspondingly.

Rudolfs[5] in 1947 reported phosphorus reductions during the course of biological treatment running as high as 75% to 90%. The average phosphate in the raw waste was 6.5 mg/l vs. 20-30 mg/l phosphate concentrations in today's waste.

Owens[4], investigating in 1953 sewage treatment plant performance in Minnesota found phosphorus removal which ranged from an average of 2%, for primary treatment plants, to an average of 23% for plants employing biological treatment. This removal was equivalent to approximately 1-2 mg/l of phosphorus. However, the lowest phosphorus contribution reported by Owens (1 5 gm/day/cap) was equal to the highest noted by Rudolfs Analysis of sewage treatment plants in the Seattle, Washington area in 1955-56 revealed comparable reductions ranging from 15% to 40% which was equivalent to 0 8 mg/l to 2.0 mg/l of phosphorus

The scattering of operating data between 15-50% phosphorus removal efficiencies by sewage treat-

ment plants is probably due to waste composition variations, mixed bacterial cultures, and microbial protoplasm with different sludge ages.

The chief factor in the design and operation of activated sludge systems for optimum nutrient reduction is the "load level" of the activated sludge, the reciprocal of which is commonly referred to as "sludge age." Figure 3 illustrates normal phos-

FIG. 3—REMOVAL OF PHOSPHATE
BY ACTIVATED SLUDGE
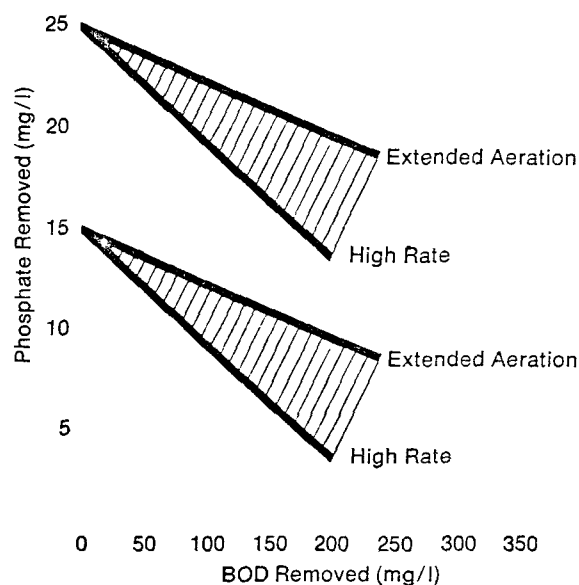
horus removal efficiencies for activated sludge sys-
ms with different sludge ages. The highest ratio of
hosphorus removed to five-day BOD removed nor-
ally occurs in the high rate activated sludge system
th the sludge age of 1-2 days Conventional acti-
ted sludge systems, working in a sludge age of 3-5
ys, have a ratio of BOD removed to phosphorus
moved of about 100 1. Extended aeration systems,
th a sludge age of over twenty days, have the
west ratio of BOD removed to phosphorus removed
a ratio of 250-500 mg BOD/1 0 mg P

One of the most recent investigations on phos-
horus removal by a different concept was con-
icted by Levin[15] and Shapiro[16] They reported
hosphorus reduction of over 80% utilizing a con-
illed activated sludge system However, one re-
w of phosphate literature[10] has pointed out that
hosphorus concentrations in the substrate used in
ir experiments were generally less than 5 mg/l.
e organic material added in these tests would
count for a phosphorus uptake of about 2 5 mg/l
syn⬤s. This is the range of removals obtained.
e slight increase in phosphorus removals above
at could have resulted from absorption. Also, only
luble phosphate was reported without regard to
total phosphate.

At San Antonio, it has been reported by Vacker,
al[17], that the phosphate removal of the Rilling
ant varies from 80-95%, averaging nearly 90% for
lengthy test period. While these removals were
easured at the Rilling portion of the San Antonio
ant, the east and west portions of the plant do not
hibit the same magnitude of efficiency. Removal
els in these portions of the plant are only 30-50%,
hough all three segments of this sewage plant
re fed from the same waste stream. The exact
echanism which produces the additional phos-
orus uptake is not established, but the following
erating conditions have been noted:

1) D.O. greater than 2 0, generally 3-5 mg/l

2) Low liquor BOD

3) High loading on MLSS, about 0.4-0.5 lbs
   BOD/lb MLSS

4) No nitrification

Although some cationic phosphorus uptake was
o noted during the test period, the significance
s not been established. Analysis of the sludge
owe⬤nsiderable quantities of iron, aluminum,
lcium, zinc, and magnesium in addition to as much
20-22% phosphate. A number of pilot plant in-
stigations, based on this approach, are being con-
ued in other parts of the United States.

Project objectives

The initial project objective was to review the exist-
ing information on phosphate removal and establish
an outline for a systems approach based on the eco-
nomical application of known phosphorus removal
parameters and also to develop new data as re-
quired After a literature study established a possible
systems approach to phosphate removal, the second
phase was initiated which included laboratory test
work necessary to prove the assumed principles of
the system. As an additional phase, commercial
scale test work was conducted where it was not
practical to conduct laboratory work or where further
study was required.

The unit operating and capital cost is considered
only as it affects the total operating and capital costs
of the sewage plant Therefore, the basic objective
was to achieve phosphate removal at the lowest
overall cost for waste treatment.

Approach

The review of literature established that chemical removal offered the only proven controllable means for phosphate removal. However, the present cost of a tertiary treatment approach was not considered practical for commercial use, except in isolated cases. Biological systems, utilizing luxury uptake, possibly hold a promise to achieve phosphate elimination at economical rates, but the degree of plant control required may make it impractical. Also, there was evidence that the method cannot be generally applied.

Of the chemical methods, lime treatment appeared the most promising, especially since the waste sludge was easily handled and it can be regenerated at a cost of 5¢-10¢ per ton CaO[18]. Furthermore, study of the phosphate removal characteristics of lime and biological systems revealed the chemical system step should be first, while the biological system scavenged the residual phosphate.

Since the prime concern when chemically treating waste water is to effectively clarify the colloidal suspension, it was decided that an approach similar to that applied to low turbidity waters should be used — that is, recirculate a portion of the settled solids. The recirculated solids provides a greater concentration of particles, thus increasing flocculating efficiency, while the recirculated calcium phosphate particles may act as a seeding agent inducing a faster rate of crystal growth.

Thus, the system developed as shown in Figure 7. Except for the recirculation of the chemical-sewage sludge floc and the use of a flocculating clarifier, the flowsheet is no different from treatment systems now employed. However, the feasibility of sludge recirculation and a system for handling the waste sludge produced had to be established. Also, test work was required to predict the probable cost of achieving phosphate removal in the manner proposed.

FIG. 4—SYSTEM CONCEPTION FOR PHOSPHATE REMOVAL
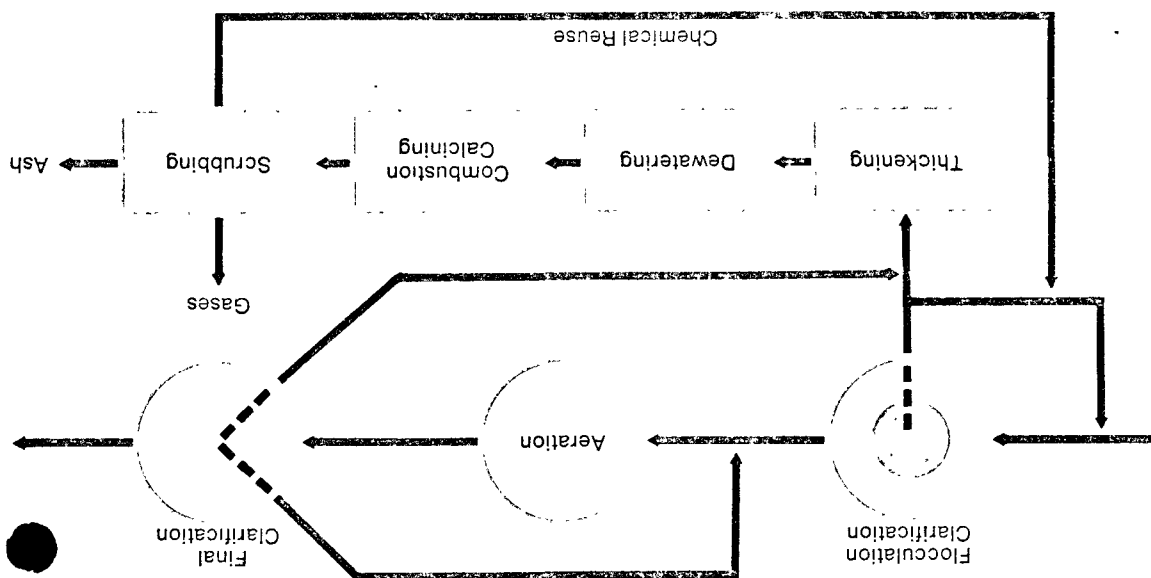
Test procedures

Phase I — Phosphate Removal

While some studies evaluating the PEP system were conducted utilizing aluminum and ferric compounds, the results were not economically promising and lime was used for the balance of all the tests. Only those results utilizing lime are herein reported.

All of the studies on phosphate removal from the raw sewage were conducted on a batch basis in the laboratory and the test work performed in the following manner:

. Sufficient lime of known available CaO was added to the raw waste samples (usually 1.0 liter) to produce a series of test vessels having a pH of 8.5-11.0. The limed sample was flocculated 15 minutes and then the mixture was filtered. The total phosphate was run on the filtrate This established an optimum curve of residual soluble phosphate vs. lime dosage and pH.

. For the recirculation studies, two liters of the raw sludge and a known dosage of lime were mixed by mechanical stirrer for 15 minutes. The lime dosage was selected to produce the desired residual phosphorus in the supernatant based on the test work of Step 1. The lime dosage was normally higher due to the poorer clarification efficiency achieved by settling vs. filtration.

. At the end of 15 minutes flocculation, the solution was transferred into a 2-liter cylinder and allowed to settle for one hour.

. After one hour of quiescent settling, the supernatant was drawn off until 200 ml remained in the cylinder. This part was classified as the underflow.

. The underflow was split and 100 ml of the underflow were transferred to the next unit which contained one liter of sewage and 100 ml of lime. This solution was mixed for 15 minutes and settled in one-liter cylinder for one hour under quiescent conditions.

. After one hour, the supernatant was drawn off until 200 ml remained, then the procedure described in Step 4 above was repeated for two or more steps which would then establish the recycle equilibrium.

Analyses were then conducted on the supernatant or total phosphate, pH, suspended solids, and, where deemed necessary, calcium, COD, and BOD data were also collected. Soluble phosphate data was only collected as an adjunct to the total phosphate determination as was alkalinity and hardness information

Concurrent tests were conducted, without the solids recirculation, to establish the chemical requirements without solids recycle. The supernatant liquor was analyzed as noted above for the recirculated tests.

Phase II — Solids Handling

After Phase I process evaluation established the character of the waste sludge to be produced by the system, laboratory tests were carried out at Dorr-Oliver's research and testing laboratories In this laboratory, a 12″ fluidized bed reactor was used to combust at 1600°F a mixture of sewage sludge and calcium carbonate sludge. (The operating temperature for calcining is 1600°F compared to the 1400°F required for sludge combustion.) The sludge lime-mud mixture was burned in the fluidized bed and the

upflowing gases strip the calcined materials and ash from the bed. The ash and the calcined material were collected by a gas cyclone and the overflow gases scrubbed in a tray unit before discharge.

Following these tests, another study was conducted at a commercial fluidized bed reactor located in Lynnwood, Washington. The lime-mud was mixed with a varying amount of sewage sludge and burned in the fluidized bed reactor. The inert solids produced by combustion were collected by a dry cyclone and wet scrubbing. The collected particles were analyzed for available calcium oxide and total calcium oxide. Figure 5 shows the testing arrangement used at Lynnwood. The lime-mud was metered into the sludge conveyor which carried the cake produced by a solid-bowl centrifuge The fluid bed reactor was designed to combust 220 lbs/hr of dry solids produced from primary clarification.

The reactor was 4'-0" I.D. (inside diameter) and had an overall height of 18'-0". The reactor is fed by a 2¾" screw feeder located in the lower portion of the bed The main air blower capacity is 400 scfm giving the reactor an input heat capacity of about 33,000 Btu/min at 4% O_2 (20% excess air) in the stack gases.

The reactor was provided with a 4" test cyclone which could recover 95% of the +2 micron particles. The cyclone was not insulated nor was the collection chamber, which consisted of a 2" section of pipe isolated by two valves to allow discharge of samples during operation. About 10% of the gas flow passed thru the cyclone while the balance was scrubbed in an impingement-type wet scrubber operating at a 10" H_2O pressure drop. The scrubbing liquor detention time was two to three minutes in the unit.

FIG. 5—COMBUSTION—CALCINING TEST ARRANGEMENTS
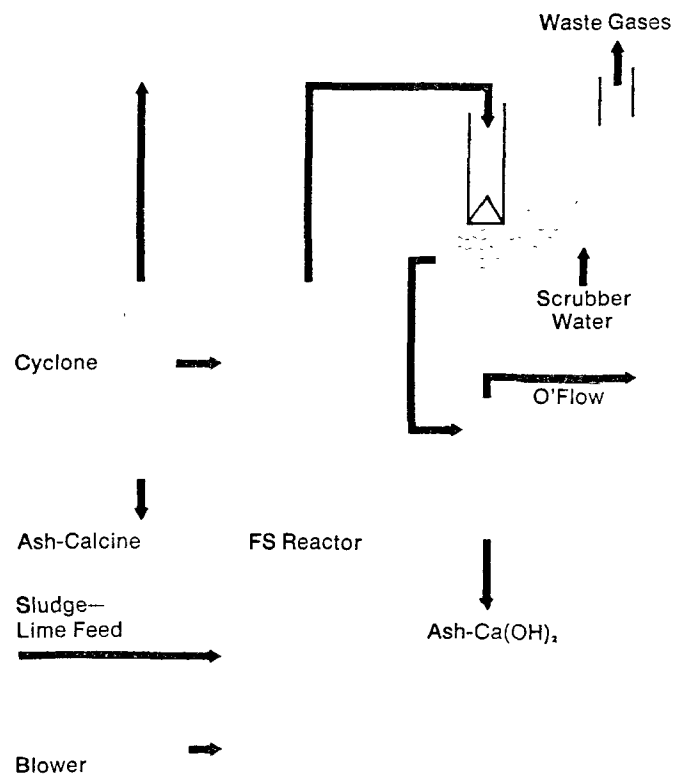
Test results

Phase I

A series of laboratory study tests were conducted at a sewage plant in Ohio where removal of phosphate was being considered In Figure 6, the phosphate removal characteristics, as a function of pH, are plotted. In this figure, both the soluble phosphate and total phosphate figures are reported. These data confirm information published by other researchers that soluble phosphate is more difficult to remove than the complex and the organic fraction.

The results are reported on pH basis since this is the easiest parameter for comparing data from different sources. In this series of tests (Figure 6), recirculation was practiced and the residual phosphate figures reported in the supernatant were obtained by laboratory clarification.

A series of studies evaluating the effects of recirculation of a portion of the settled underflow were conducted at a plant in Connecticut. This plant had relatively weak waste and a low alkalinity (about 100 ⬤ as $CaCO_3$) in the sewage A lime dosage of 70 mg/l would achieve a pH of 8 6 and reduce the total phosphate by about 65%. A dosage of 140 mg/1 of lime would achieve an 80% reduction of phosphate and produce a pH of about 9.5. At the same time, COD reductions of 65% were obtained.

In Figure 7, residual phosphate as a function of the calcium addition is shown — with and without recirculation. As this figure illustrates, the effect of recirculation was to reduce the lime addition by about 50% to achieve the same residual soluble phosphate The effect of recirculation was that comparable phosphate removal could be achieved at approximately one pH unit lower than that achieved without recirculation.

In the data of Figure 7, the information is presented on the basis of soluble total phosphate to show the effect of recirculation on the rate of calcium phosphate particle growth. The samples were filtered and the filtrate analyzed for total phosphate. The recirculation increased the rate of precipitate formation and the settling characteristics were observed to be considerably improved with recirculation.

In a third series of tests conducted at a plant in Colorado, residual total phosphate, with and without recirculation, again reflected improved clarity of the supernatant. However, at a pH of 11.0, the residual ⬤ sphate, with and without recirculation, was identical. This occured because the clarification without recirculation improved. In contrast, residual phosphate without recirculation, as shown in Figure 8 at a pH of 9 to 10 will be much higher than that experienced with recirculation.

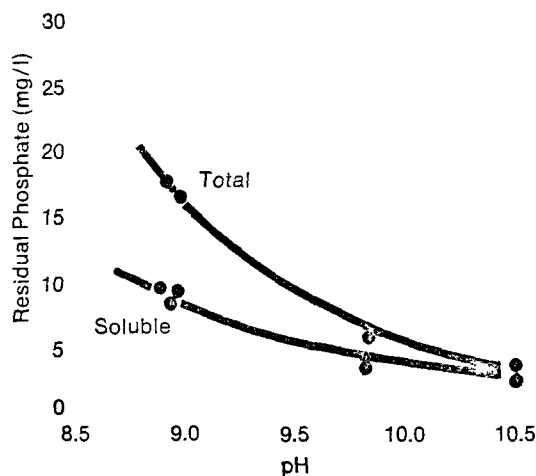FIG. 6—RESIDUAL PHOSPHATE VS pH (Plant B)

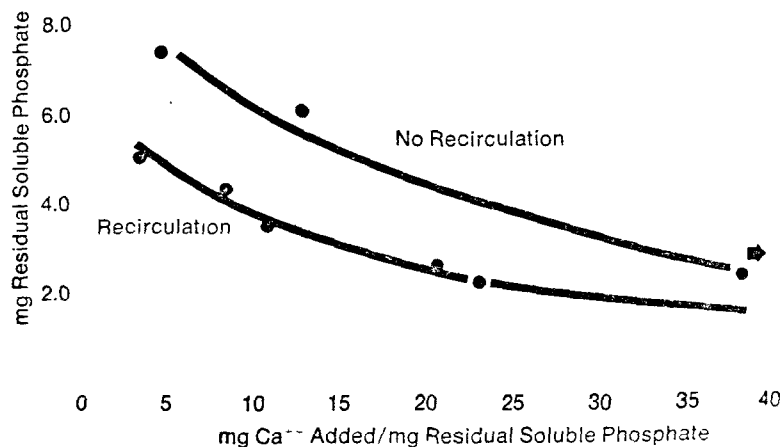FIG. 7—RESIDUAL SOLUBLE PHOSPHATE VS Ca^{-2} ADDITION (Plant A)
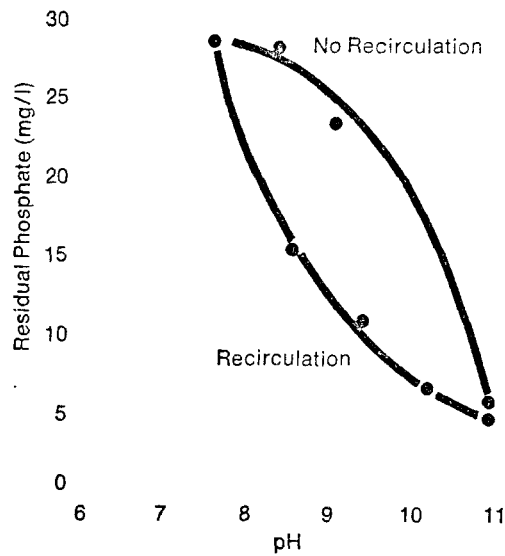
FIG. 8—RESIDUAL PHOSPHATE VS pH (Plant C)

FIG. 9—RESIDUAL COD VS pH (Plant C)
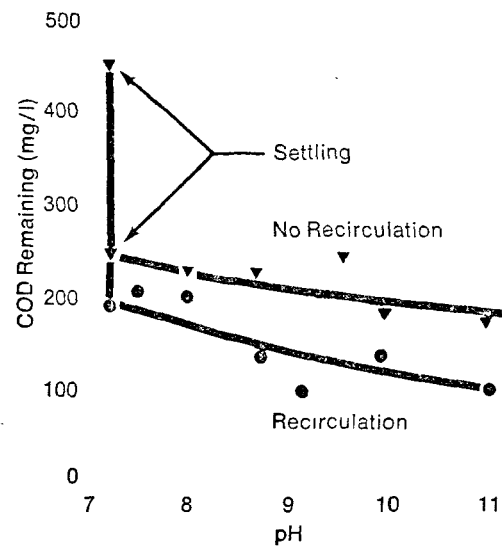

TABLE 1 — CHARACTERISTICS OF LIME TREATED RAW WASTES (PLANT C)

Sample	pH	BOD	COD	Tot. PO_4^{-3}	Total Alkalinity	Ca^{-2}	Hardness
No.		mg/l	mg/l	mg/l	mg/l $CaCO_3$	mg/l	mg/l $CaCO_3$
Raw	7.5	—	470	—	222	40	200
Settled*	7.5	186	355	28.5	—	40	—
1A-1B*	8.5	104	340	15.4	272	49	238
2B*	9.5	51	235	11.5	330	80	160
3B*	10.3	54	210	6.7	272	66	126
4A*	11.0	—	200	5.5	266	65	62

*Supernatant samples

Visual observation of the settling characteristics of the liquor anticipates these results The tests, utilizing recirculation. clarify rapidly while those samples without recirculation have considerable turbidity caused by the fine solids in suspension until a pH range of 10.5-11.0 is achieved Lab tests indicated that overflow rates of 2,000 gal/sq ft/day could provide good overflow clarity when employing recirculation.

Generally, dosing the waste to a pH of less than 10.0 will increase the calcium and alkalinity of the waste since the pH is not high enough to achieve softening. In the series of tests in Colorado, the alkalinity, calcium content and other important characteristics of the supernatant fraction were changed as shown in the table below

A corollary benefit of recirculation is an increase in the efficiency of suspended solids removal in the primary treatment step The addition of lime, combined with good flocculation and settling, increases the organic removal efficiency of the plant in the primary settling area. Figure 9 shows the residual COD characteristics of the plant in Ohio with and without recirculation.

Without recirculation, there was not a great amount of improvement in the residual COD since good clarification could not be maintained at the test conditions. However, with recirculation, it was possible to get as much as a 50% increase in COD removal at the same pH levels. Visual observation of the samples showed extremely good clarity when compared to those tests without recirculation. This better clarification efficiency, as noted by the increased COD removal, also improves removal of the calcium phosphate particles from the supernatant liquor.

The COD removal characteristics of a stronger waste, such as that previously reported in Colorado, are shown in Figure 10. Quiescent settling of the waste for one hour reduced the COD from about 475 parts to a little over 350 mg/l. With recirculation and operating at a pH of 9.5 to 10, it was possible to reduce the residual COD down to 220-250 mg/l. All the test points shown in this figure are with recirculation.

At this plant, the first series of tests showed extremely poor removals utilizing lime. The floc structure was poor and the supernatant was cloudy. Analysis of the lime used for the tests showed that was highly carbonated and contained lower available calcium oxide. These series of tests were rerun with fresh lime and checked for available CaO with standard acid.

Residual BOD data were run on the plants in Ohio

FIG. 10—RESIDUAL COD VS pH (Plant C)
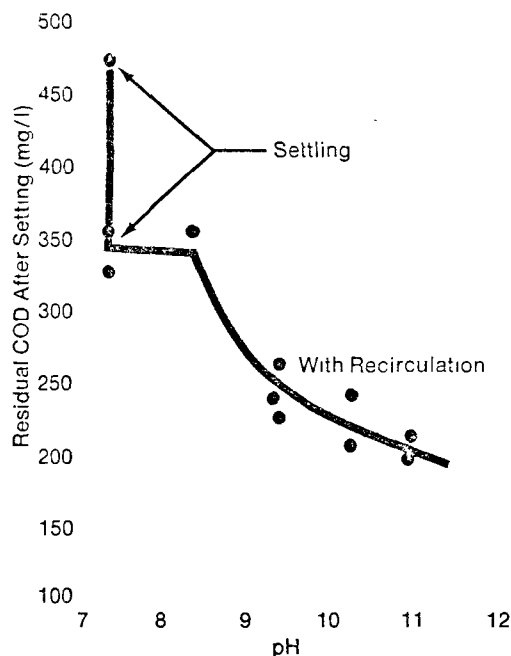
and Colorado to establish the effect of recirculation in operating at various pH's on the primary effluent BOD. In both of these plants, the raw waste strength during the tests was slightly under 200 mg/l and quiescent settling for one hour would reduce this BOD to about 110-115 mg/l or about a 38% reduction of BOD by settling. By employing the recirculation techniques described above, the results in Figure 11 show that the primary effluent BOD's can be reduced to 40-60 mg/l at a pH of 10, thus effectively reducing the load to the aeration system by 50%. With recirculation, all tests have given BOD removals across the primary step of 60-75% based on the raw BOD. In each case, the waste was predominantly domestic and had a normal soluble to total BOD ratio of 25-35%.

As reported above, the addition of calcium hydroxide to a waste, at pH's less than 10.5 to 11, will generally increase the calcium content of the waste water. The recovery of calcium in the primary sludge as a function of pH is shown in Figure 12 for two plants, Plant B and Plant C. From these results, it appears that operating at a pH of 9 5 to 10 will result in a recovery of calcium of 80-85% incorporated into the primary sludge This calcium will be com-

FIG. 11—RESIDUAL BOD VS pH

FIG. 11—RESIDUAL BOD VS pH
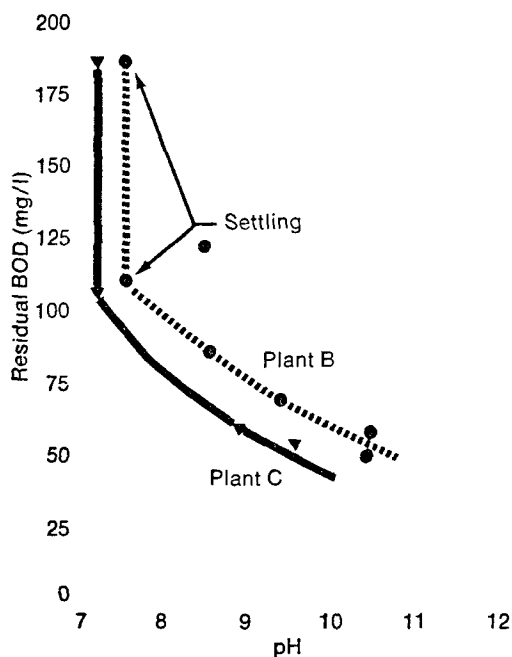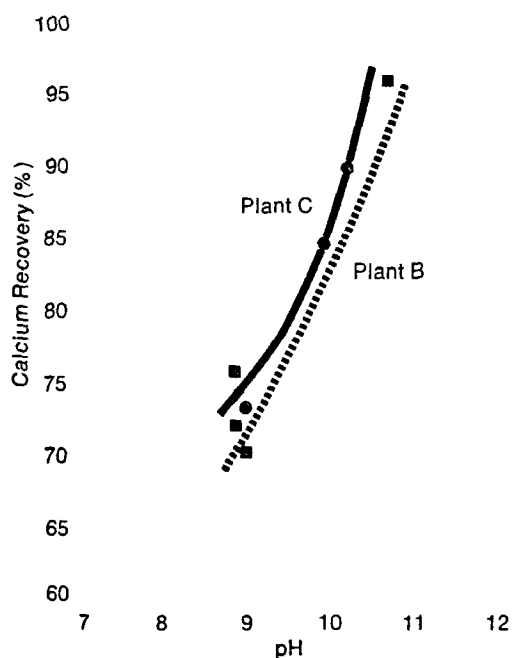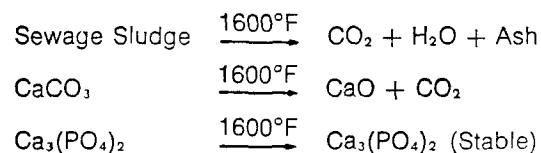
FIG. 12—CALCIUM RECOVERY VS pH

FIG. 12—CALCIUM RECOVERY VS pH

pounds of calcium phosphate and calcium carbonate. As the pH is increased above 10.5 to 11, calcium recoveries in excess of 100% will be obtained when CO_2 is used to neutralize the pH. The CO_2 source could be produced by bacteria in the secondary treatment system or from sludge combustion. The excess calcium recovery will result from softening of the waste water.

Phase II

Both calcining of lime in fluidized bed and combustion of organic sludges have been practiced for years. However, no tests have been reported on thermally treating a mixture of primary and secondary sludges with a calcium sludge. Fundamental laws governing sludge combustion would indicate that by operating at calcining temperatures the following reactions are anticipated:

$$\text{Sewage Sludge} \xrightarrow{1600°F} CO_2 + H_2O + \text{Ash}$$

$$CaCO_3 \xrightarrow{1600°F} CaO + CO_2$$

$$Ca_3(PO_4)_2 \xrightarrow{1600°F} Ca_3(PO_4)_2 \text{ (Stable)}$$

As a corollary investigation to the liquid system approach, it was necessary to establish requirements and operating conditions for combustion of the waste sludge recovered from the treatment processes. A test program was initiated whereby a sludge mixture representative of a theoretical phosphate removal plant would be incinerated in a 12″ fluidized bed reactor at Dorr-Oliver's Canal Street laboratory.

Approximately 300 lbs of filter cake from the Stamford, Connecticut treatment plant was used as a base feed. Precipitated calcium carbonate was added to the filter cake, and the moisture was adjusted to a simulated feed considered representative of a commercial installation. The characteristics of the feed were as follows:

Total Solids 29.2%
Calcium Carbonate 8.9%
Sludge Solids 20.3%
Phosphate 2.0%
Moisture 70.8%

This sludge was pumped through a progressive cavity pump into the fluidized bed reactor. After the sludge was combusted in the unit, the combustion end-products were passed through a gas cyclone and into a wet tray scrubber. The fluid bed reactor was operated with a 1600°F bed to calcine carbon-

ate. Laboratory analyses were conducted on the solids derived from the cyclone and the scrubber effluent for ash, total and available calcium and phosphorus. The test analyses showed that only 7% of the Ca^{++} recovered by the cyclone had reverted to calcium carbonate even though the cyclone temperature was only 1100-1200°F. All of the calcium in the wet scrubber was in the carbonate form since the water was recirculated and was being contacted with the waste gases containing 16-18% CO_2.

The commercial scale combustion facilities at Lynnwood, Washington were deficient in that the dry cyclone was not insulated. The sample collection period was approximately 10 minutes and the temperature was not in excess of 500°F with 18% or more CO_2 present in the stack gas. The temperature in the sample column was sufficiently low to allow condensation during some samples.

Composite samples of the feed lime-mud, sewage sludge, and cyclone underflow collected during the tests showed the analyses in Table II.

⬤he cyclone, even though it could remove the majority of particles down to one micron, exhibited a haziness in the waste gases indicating that a significant portion of the ashed solids was finer than one micron. Wet scrubbing was necessary to clean the gases to meet air pollution requirements.

The percent available lime in the cyclone under-

flows varied from 79.6% to 90%. The 5 6% available lime was that which was found in the ash of the feed sludge without lime added to the reactor Available lime in the overflow stream from the DorrClone® varied from 43.7% to 70%. (Table III)

It was necessary to feed a small amount of sludge feed to the reactor at all times. The lime-mud was sufficiently dry to cause binding of the feed screw unless the sewage sludge was added The weight of sludge to the reactor was 25-75 lbs. total solids per hour during the majority of the tests The reactor was fed 420-630 lbs/hr of lime-mud utilizing 380-400 scfm air at 1.0% to 3.0% O_2.

TABLE II — FEED/PRODUCT ANALYSES

	Total CaO	Available CaO	Volatiles
	%	%	%
Sewage	1.0	—	82
Lime Mud	36.8	0.74	—
Cyclone U'Flow	59.0	47.2	—

TABLE III — ANALYSES OF PRODUCTS

				Available CaO/Total CaO		
Time	Sample No.	Temp — °F		Cyclone	Scrubber U'F	D/C**O'F
		Bed	FB*	%	%	%
10:50	1	1680	1650	5.6	—	—
11:20	2	1560	1670	86.2	64.5	70.0
13:05	3	1610	1780	87.5	62.8	59.0
13:25	4	1490	1790	83.5	58.4	50.0
16:35	5	1640	1800	79.6	64.8	61.6
16:42	6	1600	1800	—	54.0	56.0
16:46	7	1580	1800	90.0	60.0	58.2
17:04	8	1480	1780	—	—	56.8
17:34	9	1400	1760	—	—	56.2
18:11	10	1320	1740	—	—	43.7

*Feedboard space of reactor
**Ash cyclone (DorrClone)

Discussion of results

The test results and data analyses indicated that about 90% phosphate removal was easily obtained at a cost commensurate with present capital and operating costs of activated sludge treatment plants. The operating costs increase when removals of 95% and higher are required. However, for many receiving streams, high degrees of phosphate removal are not justified, and particularly so, if correspondingly high organic carbon removals are not simultaneously obtained.

The carbonaceous material entering a receiving water is broken down by bacteria to its lowest oxidized state, CO_2 and H_2O, or in the benthal deposits, the reduced state of CO_2 and CH_4. The CO_2 produced by the bacteria are utilized by the algae as a carbon source. Futhermore, benthal deposits act as a reservoir of nutrients which becomes available during the spring turnover of lakes.

There is a symbiotic relationship between the bacteria and algae. Effluent requirements of BOD, total suspended solids, and phosphate should be considered in this light. The State of Pennsylvania recently took steps which were consistent with the above approach in requiring treatment plants in a specific watershed area to produce effluents not greater than 4 mg/l BOD and 0.2 mg/l P.

In the subsequent discussion of removing phosphate with calcium, we will refer to it as calcium phosphate. It is not practical to differentiate between the many possible forms of insoluble phosphate that might be present. The chemistry of phosphate is not precise, and it is particularly difficult in the heterogeneous mixture of ions present in sewage. When calcium phosphate waste sludges are calculated, they will be considered on the basis of tri-calcium phosphate, although other forms may be present.

Phase I

Laboratory data established that the chemical treatment of the raw waste is an efficient means of removing the bulk of the phosphate in the sewage. Not only is the majority of the phosphate removed, but also substantial increases in BOD removal are achieved. In the tertiary treatment system for phosphate removal, a proportion of the phosphate is removed by primary and activated sludge. After that, it is necessary to add lime or another coagulant to reduce the

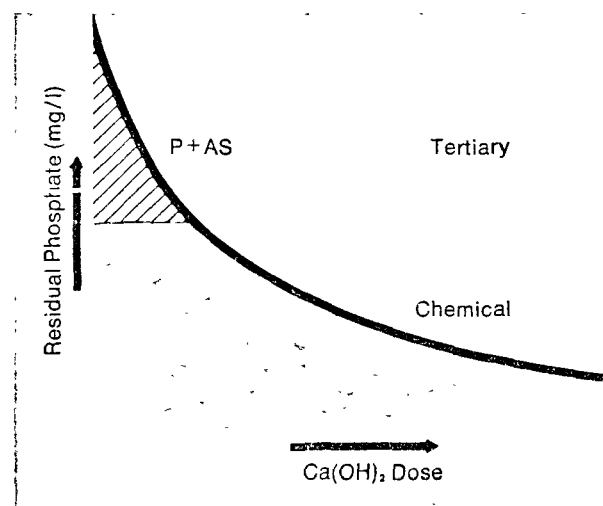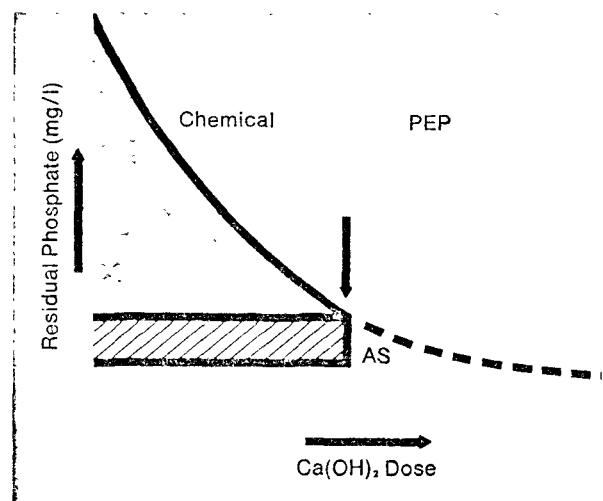FIG. 13—TERTIARY TREATMENT CHEMICAL REQUIREM

FIG. 14—PEP SYSTEM CHEMICAL REQUIREMENTS

phosphate to the desired level. The chemical requirements for this procedure is quite high In addition, substantial equipment requirements are required. Graphical representation of the tertiary treatment chemical efficiency is shown in Figure 13 which illustrates a monomolecular relationship of residual phosphate to lime dosage.

In contrast, if the chemical treatment is practiced first and the residual scavanged by biological treatment, we get a removal characteristic as shown in Figure 14. The test results establish that this method, in addition to the recirculation of a substantial portion of the clarifier underflow after flocculation, can reduce lime requirements by as much as 60-75%.

The higher removals of phosphate at lower pH's are most likely due to the recirculation of calcium phosphate nuclei. This nuclei would act as a seed for increasing the rate of phosphate removal from solution. This may also account for the excellent settling characteristics of the raw waste stream after mixed with the recycle solids for short contact times. The nuclei theory is reinforced by Clesceri[11] who reported that colloidal chemical studies showed that at a pH of 11, the pre-dominant calcium phosphate compound is hydroxylapatite which may be mainly micro-crystalline and therefore difficult to settle.

Using lime, the detention time in the Flocculator greatly influences the efficiency of phosphate removal. Detention times of one hour have been required for many applications. It would appear that considerable time is required to build the calcium phosphate particle to such a size that it will settle readily. With the application of recirculation, a large quantity of previously precipitated calcium phosphate particles is maintained in circulation. The effect of this recycle is to hasten the growth of the calcium phosphate particles by seeding effect. This seeding allows reducing the detention time down to not more than 15 minutes while achieving the results noted above.

The results indicated that pH's of 9 5 to 10 in the Flocculator-clarifier unit will be sufficient to maintain the overall removal of about 90%. The system may be controlled at the pH necessary to achieve the desired phosphate residual This may be done by monitoring the pH in the Flocculator-clarifier and adding the makeup calcium hydroxide as required. A curve of residual phosphate versus pH can be established for a specific plant and used as a control parameter for monitoring lime addition It would also be possible to use total phosphate analysis as the control parameter, employing a continuous analyzer.

Phase II

The combustion testwork, operating at calcining temperatures, proved that the sludge mixture is easily handled in the present fluidized bed conception. It would appear that the principal consideration for producing a completely calcined product is to maintain the bed within known temperature limits required for calcination.

The testwork established that if the calcium oxide is quickly removed from the gas stream without severe cooling, very little recarbonation will occur. There are a number of devices on the market which can remove the calcined material from the stack gases, either in dry or wet form. Actually, calcination of a lime-mud sewage sludge mixture is not unusual, as many vacuum filter cakes may have 20-25% calcium carbonate present from the lime used to condition the sludge.

Evaluation of the system

The resulting sludge from the PEP system will be significantly different than that derived from a conventional treatment plant or a tertiary treatment system. The waste stream will contain much less activated sludge. Table IV shows the relative sludge quantities from a million gallon plant employing a conventional primary plus activated sludge system with an additional tertiary treatment and also a comparison with the PEP system

While the PEP system will have considerably more total sludge than the conventional system, the sludge characteristics are much different. Most important, there is much less activated sludge The tertiary treatment system will have about two to three times as much sludge as the PEP system.

The operating costs for sludge handling vary depending upon the type of sludge. Table V shows typical operating costs per ton for the various types of sludge derived from sewage treatment Primary sludge, for example, can be dewatered by vacuum filtration or centrifugation for about $3-$5 per ton. The sludge can be dewatered to a concentration that is thermally self-sufficient Conversely, activated sludge by itself or mixed with the primary sludge will cost $15-$25 per ton for chemicals for dewatering[19] [20] and the fuel costs will be about $12-$15 Calcium carbonate particles being quite heavy are easily dewatered and require $4-$5 of fuel per ton to calcine.

It can be considered that the operating cost for various sludge mixtures will be dependent upon their proportionate amount This is generally true for the sludges produced in domestic waste treatment. The cost will be $30-$35 per ton of activated sludge to dewater and burn and if this sludge quantity is reduced, it will result in a significant savings. In the PEP flowsheet, it is possible to reduce this sludge quantity by as much as 60%, thus achieving a significant savings in the operating costs for handling activated sludge.

Table VI shows the chemical costs of phosphate removal, reported by Nesbitt[10], and also that for the PEP System. In this comparison, the Dorr-Oliver lime dosage was established at 200 mg/l, even though high efficiency has been maintained at much lower dosages. These costs consider that the attri-

TABLE IV — WASTE SLUDGE PER MGD

Sludge	Conventional	Conventional and Tertiary	PEP
	lb/day	lb/day	lb/day
Primary	1250	1250	1780
Secondary	630	630	250
$CaCO_3, Ca_3(PO_4)_2$	—	5100[1]	1810[2]
	1880	6980	3840

(1) 450 mg/l $Ca(OH)_2$ — 100% Recovery Ca^{++}
(2) 200 mg/l $Ca(OH)_2$ — 80% Recovery Ca^{++}

TABLE V — OPERATING COSTS FOR COMBUSTION

	Oper. Cost — $/Ton	
Sludge	Dewater	Fuel
Primary	3-5	0
Act. Sludge	15-25	12-15
$CaCO_3$	0	4-5

TABLE VI — CHEMICAL COSTS* OF PHOSPHATE REMOVAL

Coagulant	Dose (mg/l)	Cost $/MG-Yr
$Ca(OH)_2$	450	10,200
$Al_2(SO_4)_3 \cdot 18 H_2O$	225	17,400
$FeCl_3$	100	25,800
$Fe_2(SO_4)_3$	150	9,350
PEP — Lime	200	3,390

*Based on 8 mg/l P in raw sewage.

tion loss of calcium is 40% and that the lime cost as 100% calcium oxide is $20 per ton. Of the 200 mg/l dosage, 120 mg/l are provided by recovery through a combustion-calcining system

In comparing the costs, Table VI shows that the chemical cost of the PEP system is $3390 per million gallons per year. However, this cost is not representative of the actual operational cost of the overall plant since there are substantial reductions in the quantity of activated sludge to be handled. In addition, since the BOD is reduced by 60% or more across the primary system, there is correspondingly less power required to operate the aeration system The net effect of this reduction in power and elimination of waste activated sludge on the overall operating cost of the plant is shown in Table VII.

Table VII shows that the additional cost of calcium hydroxide makeup at 40% attrition (including the cost of reburning the calcium carbonate sludge) is less than the savings resulting from the elimination of 65 tons of activated sludge plus the savings in aeration power. The cost of $3390 per million gallons per year was more than offset by a savings of $4560 per million gallons per year in operating costs for the waste sludge handling and power for activated sludge system. If there was no lime recovery practiced, the cost would be $4550 a year or equivalent to the savings attributed to calcium hydroxide treatment in the primary stage.

Lime recovery

The pilot plant and the commercial tests indicated that calcining of lime while burning sludge is quite practical The end product from combustion may be collected by either dry or wet scrubbers with low liquor detention times and then after slaking, the lime is extracted from the ash The ash is wasted from the system while the calcium hydroxide is recycled to the primary treatment stage It is necessary to operate the combustion unit at a slightly higher temperature than that required for complete deodorizing. However, the savings attributed to the costs of handling activated sludge makes it economical to burn this mixture and recover the lime particularly for the larger waste treatment plants Due to the elimination of much of the hydrous activated sludge, the dewatering and combustion equipment requirements will be no more than that required for normal conventional treatment plants

The decision to recover lime from ash would depend upon the size of the treatment plant, cost of purchased lime, and ash disposal considerations. A typical plant could have the following ash quantities produced for disposal:

TABLE VII — NET OPERATING COST FOR PEP SYSTEM

	Tons/MG-Yr	$/ MG -Yr. Cost	$/ MG -Yr. Savings
Ca(OH)₂ Makeup	122	1830	—
Ca(OH)₂ Recycle	182	1560	—
Activated Sludge	—65	—	2370
Power at 1½¢/kwh		—	2190
		3390	4560

Cost w/o Lime Recovery $4550

TABLE VIII — ASH QUANTITIES FOR ULTIMATE DISPOSAL

	Conventional lbs/MGD	PEP System — lb/MGD W/Reuse	PEP System — lb/MGD W/o Reuse
Sludge Ash	400	400	400
Ca₃(PO₄)₂	—	320	320
CaCO₃	—	300	1490
Total — lbs/MGD	400	1020	2210
— cu ft/MGD	4	10.2	22.1

Application of the results

The liquid treatment portion of the phosphorus removal system is shown in Figure 15. In this system, the raw waste is flocculated and clarified preferably in an integral unit. To the raw sewage, clarifier underflow solids are recycled to maintain the suspended solids in the mixture fed to the flocculator of about 500-2000 mg/l. Lime is added as required to maintain the pH at the desired point. The lime addition is automatically controlled and would fluctuate depending upon raw sewage flow, the alkalinity, and the required phosphorus removal.

FIG. 15—PEP TREATMENT SYSTEM
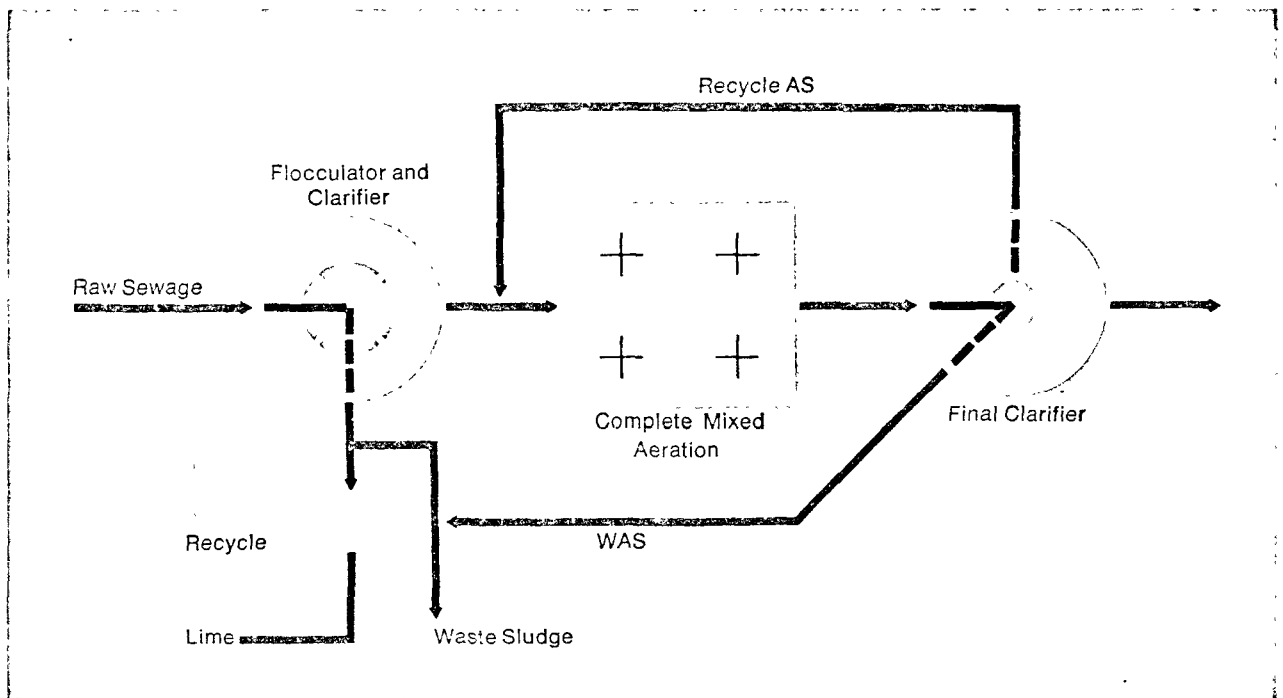
Nitrogen removal

The raw sewage from which more than 80% of the phosphate and suspended solids and about 60-70% BOD have been removed would be discharged into an aeration system employing the complete-mix activated sludge concept This system would allow the addition of primary clarifier effluent to the aeration tank without a pH correction The homogeneity of this activated sludge system would eliminate much of the inhibitory effect of a relatively high pH In a plug-flow system, the high pH could have a serious effect on the biological culture The advantages of the complete-mix aeration system over the plug-flow system have been described by other investigators in the sanitary field

There will be an inherent pH adjustment in the aeration basin when the feed has a pH of 9 5 to 10 5 This will come from the liberation of the CO_2 by the activated sludge culture during metabolism of the remaining BOD Depending upon the operating parameters of the aerobic system, the activated sludge can provide 50-100% of the necessary pH adjustment thru the production of CO_2 by the microbic cells. This adjustment will produce a pH below 9 0 If it is necessary to adjust the pH to a lower value than that achieved by natural CO_2 production, then acid or stack gases from the combustion unit can be utilized.

Since the BOD added to the aeration basin will be in the neighborhood of 50% of that normally present, the tankage requirements may be correspondently reduced In fact, this reduction must be included as it is desirous to operate the activated sludge culture at a high loading utilizing a high MLSS. It is quite possible that the effluent BOD's will reflect the large reduction in the influent BOD The significant consequence of this lower BOD to the aeration basin is a reduction in the quantity of biological sludge which must be wasted It must be expected that there will be some calcium carbonate precipitation in the aeration basin This will be recycled with the normal activated sludge return to the aeration basin No deleterious effect from this is anticipated, and it should enhance settling in the final clarifier

The final clarifier should be of the rapid sludge return type. Removing the activated sludge from the final clarifier quickly will minimize the amount of leaching of phosphate into the liquor from the activated sludge cell. Phase separation of the activated sludge may make it possible to separate the calcium carbonate precipitation by wastage of the sludge near the influent portion of the clarifier. The waste activated sludge is discharged to the primary treatment portion or directly to the solids handling area

If nitrogen removal is required, it will be necessary to raise the pH in the primary clarifier effluent to 10 5 to 11 At this pH, the ammonia nitrogen can be air stripped from the waste This can be most economically conducted in the manner described by Culp[6] which is to pump the waste into an ammonia stripping tower wherein air is added in large quantities countercurrent to its sewage flow. The reported quantities of air have been as high as 500-700 scfm per gallon of sewage to achieve removals of 90-95% of the ammonia when the system is operating as a tertiary step

The characteristic removal of ammonia by air stripping will be similar to that of the removal of phosphate by lime. The rate of ammonia removal is a function of the partial pressure of the ammonia in the waste and in the air Since there will be a downstream requirement for ammonia as a nutrient by the activated sludge system, a higher residual can be left in the primary effluent This will greatly reduce the amount of air required to strip the ammonia from the waste.

When nitrogen removal is practiced, it will probably be necessary to neutralize the effluent by the addition of CO_2 or another acid source. One CO_2 source could be the off-gases from the combustion unit which is destroying the waste organics However, Sawyer and Buzzell[12] were able to add the primary effluent at a pH of 11 to an aeration chamber without severe inhibitory effects, but there was a considerable buildup of calcium carbonate in the recycle sludge stream

Sludge disposal

The sludge handling flowsheet for the PEP System is shown in Figure 16. Sludge is wasted from the primary system at a rate necessary to maintain the desired recirculated solids concentration. As the waste rate will vary, a varying quantity of dilution water must be provided. The diluted primary underflow is pumped through a low pressure hydrocyclone (DorrClone) to remove all of the +150 mesh grit.

The waste primary and secondary sludge is thickened by gravity Thickener area requirements for the PEP approach are the same as that required for a conventional plant of similar capacity. This is due to the large reduction in the waste activated sludge which is difficult to thicken and dewater. Dewatering of the predominately primary sludge will be easily accomplished in a "long bowl" solid bowl

FIG. 16—PEP SOLIDS HANDLING SYSTEM

conveyor centrifuge (MercoBowl®) Chemical costs will be lower and final cake concentrations much higher due to the presence of $CaCO_3$ and the reduced amount of activated sludge

Combustion of the sludge mixture can be efficiently carried out in a fluid-bed reactor (FluoSolids Reactor) where critical temperatures can effectively be controlled. Waste gases carrying the sludge ash and the dehumidified to reduce the water vapor plume produced by the saturated gases.

The scrubber water containing ash and the slaked lime would be transferred to a thickener where sufficient dilution water (from the thickener overflow) would be added to dissolve the calcium hydroxide suspension. The overflow would be returned to process and additional lime added as required to maintain the pH.

The ash and calcium phosphate particles are settled and pumped through a small hydrocyclone (DorrClone) to classify out the solids. The overflow is returned to the thickener feed while the phosphate h is dewatered on a mechanical classifier.

Summary and conclusions

1. Laboratory tests were conducted on a complementary liquid and sludge handling treatment system which involved chemical removal of phosphate and a greatly increased BOD removal followed by activated sludge to reduce the balance of the phosphate and BOD to the desired level. The PEP System can reduce the operating costs associated with phosphate removal to a level equivalent to that required for conventional treatment considering the overall operating cost for waste treatment.

2. Both laboratory and commercial tests established the fact that combustion of lime mud with sewage sludge is practical and that calcium oxide can be recovered from the combustion ash The combustion of sludge must be conducted at a temperature sufficient for calcination of the calcium carbonate. The recovery of the calcium carbonate as calcium oxide can be conducted at one-half the cost of purchased lime and reduces the problem of disposing of large quantities of chemical precipitate

3. The PEP System design is similar to that of a conventional plant. Allowing for a 40-50% reduction in the aeration tank volume, it is possible that the PEP System plant can be built for a cost comparable to a conventional activated sludge plant employing incineration for disposal of the waste sludge.

4. Higher removal efficiency of phosphate combined with ammonia nitrogen removal can be achieved with the same basic approach However, operating and capital cost will exceed conventional treatment costs.

5. Because of the high pH in the raw waste, an additional benefit of the PEP System will be the elimination of sulfide odors commonly found in the primary treatment portion of sewage plants.

6. The complementary approach of phosphate removal, using the most economical combination of known phosphorus removal mechanisms, now provides the engineer with an important tool to achieve significant levels of phosphate reduction without incurring penalties of increased operating cost, capital cost and larger land requirements.

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PHOSPHATE REMOVAL
BY
CHEMICAL PRECIPITATION

J. H. DUFF

R. DVORIN

E. Salem

Graver Water Conditioning Co.

INTRODUCTION

Properly treated sewage effluent can be a valuable source of
industrial process water, cooling water and boiler feedwater. This is
particularly true in the chronically water short-industrial areas in
the Southwest and West. (1) In 1946, the first of many Graver plants
designed to recover industrial water from treated sewage went on stream
to supply a Kansas oil company with cooling water. Experience at this
installation, and others, indicated to users that phosphate reduction
to low levels was necessary to avoid phosphate scale deposition on heat
exchange surfaces. As a result of this need, industrial technology
rapidly developed to the point where phosphate reduction to levels of less
than 1 ppm PO_4 as $CaCO_3$ (0.6 ppm as PO_4, 0.2 ppm as P) became routine.

This industrial technology may now be applied to waste treatment
facilities where phosphate reduction is necessary as part of an overall
effluent upgrading program. Principal equipment and operating costs are
available, based on many long term large scale industrial installations.

PRINCIPLES OF PHOSPHATE REDUCTION

Phosphates may be removed from solutions by precipitation. Some
of the possible reactions are shown in Figure 1. Commercial factors limit
the chemical reagents applicable. The principal reagents used are lime (CaO)
and alum $(Al_2(SO_4)_3 \cdot 18 H_2O$. Figure 2 is based on the results of labora-
tory jar tests run on sewage and industrial phosphate bearing wastes. The
tests were run at ambient temperatures using standard jar test procedures. (2)

In virtually all cases, both poly and ortho phosphates were present in

the samples. The removal efficiency indicated is based on total phosphate.

The phosphate removal by calcium precipitation in Figure 2 is based on tests in the pH range of 9.5 to 11. (1 ppm calcium as Ca is equivalent to 0.6 ppm of commercial 93% quick lime-CaO).

It would appear that this phosphate removal proceeds on an equivalent basis. This presumes the calcium demand (aside from phosphate) of the system is satisfied. The calcium demand is essentially that required for reactions with constituents other than phosphate and to leave a residual of 75-140 ppm calcium (as $CaCO_3$) in the final treated water. It should be pointed out that in the treatment of municipal sewage and some industrial wastes, calcium carbonate precipitation is inhibited by as yet unidentified materials.

Phosphate removal, using aluminum or iron salts, appears to be less efficient than lime on an equivalent basis. The curves on Figure 2 for aluminum and iron salts are based on a pH range of 6.5 to 7.5. The degree of phosphate removal affects the chemical efficiency, that is, the more complete phosphate removal required, the more equivalents of precipitant per equivalent of phosphate. (1 ppm Al as $CaCO_3$ = 0.2 ppm of alum ... $Al_2(SO_4)_3 \cdot 18 H_2O$).

Figure 2 is based on analytical results using filtered samples. In both the calcium and the aluminum or iron systems, some coagulant is required, since simple precipitation of the phosphates does not insure a low total phosphate residual unless the suspended solids concentration is low.

PROCESS DESIGN

Process design must insure maximum chemical efficiency, minimum equipment and operating costs, and low effluent turbidity in the final treated water. Requirements of the ideal system are shown in Figure 3.

The chemical treatment used must produce the required treated water characteristics. The end use of the water determines these characteristics (analysis); such as, pH, alkalinity, phosphate, hardness, suspended solids. For example, underground disposal requires a stable water low in suspended solids, while industrial uses generally require low dissolved and suspended solids. Knowledge of the end use requirement is needed to choose the lowest cost chemical treatment (3, 4).

To utilize most effectively the chemical treatment applied, a proper environment must be provided. This includes rapid uniform dispersion of all reagents into the system in their proper order. Previously formed precipitates which enhance the rapid growth of new precipitates when provided with gentle agitation, should be present. Sufficient time must be provided in the system for the reactions to go to completion and for maximum particle growth.

Equipment scaling is controlled by maintaining high concentrations of previously formed precipitates in circulation so that new precipitates form on the surfaces of existing ones. The absence of sufficient quantities of previously formed precipitates can severely hamper equipment performance. Precipitates must be separated from the treated water. The settled sludge volume is minimized by mechanical thickening.

OPERATING EXPERIENCE

The Graver Water Conditioning Co. has supplied equipment for the treatment of sewage plant effluent at a number of locations. The designs are based principally on water re-use for boiler feedwater, cooling tower makeup, and general plant water. (2) One of these plants, Figure 4, treats Amarillo, Texas, municipal waste water principally for cooling tower makeup.

The first of the four Reactivators ^{R} was installed at this electric generating utility in 1958. The two 56' diameter and two 70' diameter Reactivators have a combined rating of 13 mgd. Lime treatment of the sewage plant effluent was chosen to provide a dependable supply of cooling tower makeup (5).

Figure 5 shows the functional design of the Graver Reactivator. This high rate, controlled recirculation, solids contact clarifier is ideally suited for phosphate removal as the design incorporates all features necessary. These include:

1. Rapid mixing and recirculation zone for flash mixing of previously formed precipitates for treatment chemicals and incoming raw water.

2. Slow mixing and floc formation in circulation zone for maximum precipitate growth and adsorption.

3. Quiescent settling zone provides for separation of the precipitates from the upflowing treated water.

4. Sludge collection zone.

5. Positive sludge scraper.

6. Final thickening before sludge discharge.

Figure 6 shows typical operating results obtained at this installation. Treated water requirements are being met by almost complete phosphate removal. For cooling water purposes, silica reduction is desirable and is being obtained. The chemical treatment demand varies at this installation as does the inlet water composition. Lime demand is estimated as the sum of the following:

Magnesium reduction

Bicarbonates

Free Carbon Dioxide

Coagulant

Phosphate reduction

Excess hydroxide

Operating results obtained are in excellent agreement with this method of determination. Under the conditions shown on Figure 6, approximately 2.5 lbs. of lime (CaO) and 0.25 lbs. of alum are used per 1000 gallons of inlet water.

Phosphate Reduction Using Alum. The 40' diameter Reactivator operating at the County of Nassau, Water Renovation Project uses alum for phosphate removal and as a coagulant. The purpose of this plant is to treat water for re-injection into the ground to block sea water infiltration. The flow diagram is shown in Figure 7. The treated water quality requirements are:

1. Chemical constituents not to exceed U.S.P.H.S. Standards for drinking water quality.

2. Turbidity not to exceed 1 JTU.

3. COD not to exceed 5 mg/l.

Treated water quality requirements are met by coagulation and phosphate precipitation in the Reactivator followed by polishing for suspended solids removal in the dual media air scour filters. Final polishing for organic removal is carried out in the granular activated carbon adsorbers.

An important feature of the Reactivator at this installation is the recently developed hydraulic recirculation system. Recirculation systems as used in Reactivators as shown in Figures 4 and 5, require electrical power to drive the mechanical circulation system. The hydraulic recirculator utilizes the kinetic energy in the incoming raw water for the recirculation of previously formed precipitates and the mixing of chemicals and raw water. This equipment can be operated with either the electrically powered mechanical recirculator or with the hydraulic recirculation system.

Figure 8 shows operating results obtained at this installation with the hydraulic recirculator in service. At the time of these tests, only alum was being fed to the Reactivator although jar tests indicated lower turbidity levels would be obtained with a polyelectrolyte. To obtain the results shown, 1.7 lbs. of alum per 1000 gallons was used.

It may be noted that the total phosphate concentration of the Reactivator effluent was 6.8 ppm as calcium carbonate, while the turbidity was 6 JTU. Virtually complete removal of the turbidity by filtration re-sulted in virtually complete removal of this residual total phosphate, thus

indicating that the major part of the phosphate leaving the Reactivator was suspended rather than dissolved material. We believe that with improved chemical treatment using a polyelectrolyte, the Reactivator effluent turbidity and total phosphate level would be significantly reduced.

PLANT DESIGN

There is no practical limit to the size plant that can be provided for phosphate removal from municipal or industrial waste water. Clarification units, such as the Graver Reactivator, are currently in operation in sizes ranging from 8' up to and including 175' diameter. Commercially available equipment provides an individual unit capacity over the entire range of 0.1 to 30 mgd. Larger flows, can be handled by multiple unit installations. Virtually any phosphate level can be reduced to any desired level down to a practical limit of about 1 ppm as $CaCO_3$ (0.6 ppm PO_4, 0.2 ppm as PO_4 as P).

Costs. Figure 9 is an example of a 10 mgd phosphate removal plant using lime treatment. For the purposes of this example, the sewage plant effluent analysis given on Figure 6 was used. The chemical dosage levels are based on 80% phosphate reduction, that is, reduction, to a level of 8 ppm or less, as $CaCO_3$.

Sludge Handling. Under the conditions of this example, approximately 3 lbs. of dry weight solids are precipitated from each 1000 gallons of water treated. For a 10 mgd plant this is equivalent to approximately 30,000 lbs. (15 tons) of dry solids for disposal per day. For design

7

purposes, a blowdown sludge concentration of 5% should be used, although concentrations obtained by operating equipment is often in the range of 10-15% solids by weight. Assuming an underflow concentration of 5% by weight, this would be equivalent to a volume of about 70,000 gallons per day. Sludge de-watering equipment would be required to reduce this. Although centrifugation may be used, vacuum filtration will reduce the moisture content so that the sludge would have a maximum moisture concentration of 50% by weight. This would result in a maximum of 60,000 lbs. (30 tons) of de-watered sludge, equivalent to a daily volume of approximately 20 cu. yards.

SUMMARY

Phosphates are removed by chemical precipitation using standard commercially available equipment. Raw water composition and treated water requirements are the deciding factors in choosing between lime and alum treatment.

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FIGURE 1

PRECIPITATION OF PHOSPHATES

$Al_2(SO_4)_3 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + 2AlPO_4$

$3BaCO_3 + 2Na_3PO_4 \longrightarrow 3Na_2CO_3 + Ba_3(PO_4)_2$

$3Ca(OH)_2 + 2Na_3PO_4 \longrightarrow 6NaOH + Ca_3(PO_4)_2$

$3CuSO_4 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + Cu_3(PO_4)_2$

$Fe_2(SO_4)_3 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + 2FePO_4$

$3MgSO_4 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + Mg_3(PO_4)_2$

$2MgSO_4 + (NH_4)_2SO_4 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + 2MgNH_4PO_4$

$3NiSO_4 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + Ni_3(PO_4)_2$

$3ZnSO_4 + 2Na_3PO_4 \longrightarrow 3Na_2SO_4 + Zn_3(PO_4)_2$
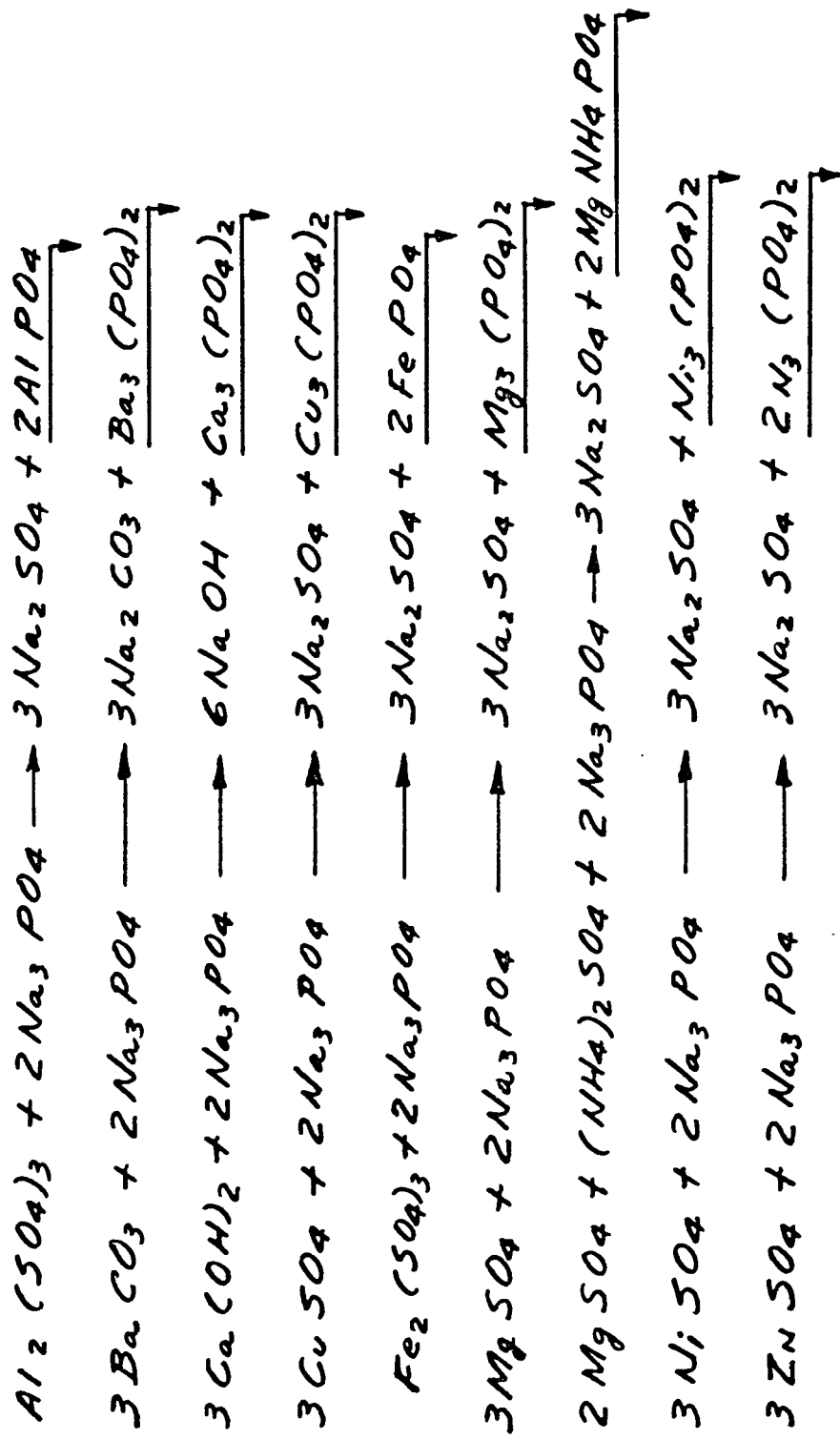
PRECIPITANT REQUIREMENTS

INITIAL PHOSPHATE (PPM CaCO₃)

PHOSPHATE REMOVAL (Ca)

80% PHOSPHATE REMOVAL (Fe-Al)

90% PHOSPHATE REMOVAL (Fe-Al)

95% PHOSPHATE REMOVAL (Fe-Al)

PRECIPITANT DOSAGE - (PPM CaCO₃)
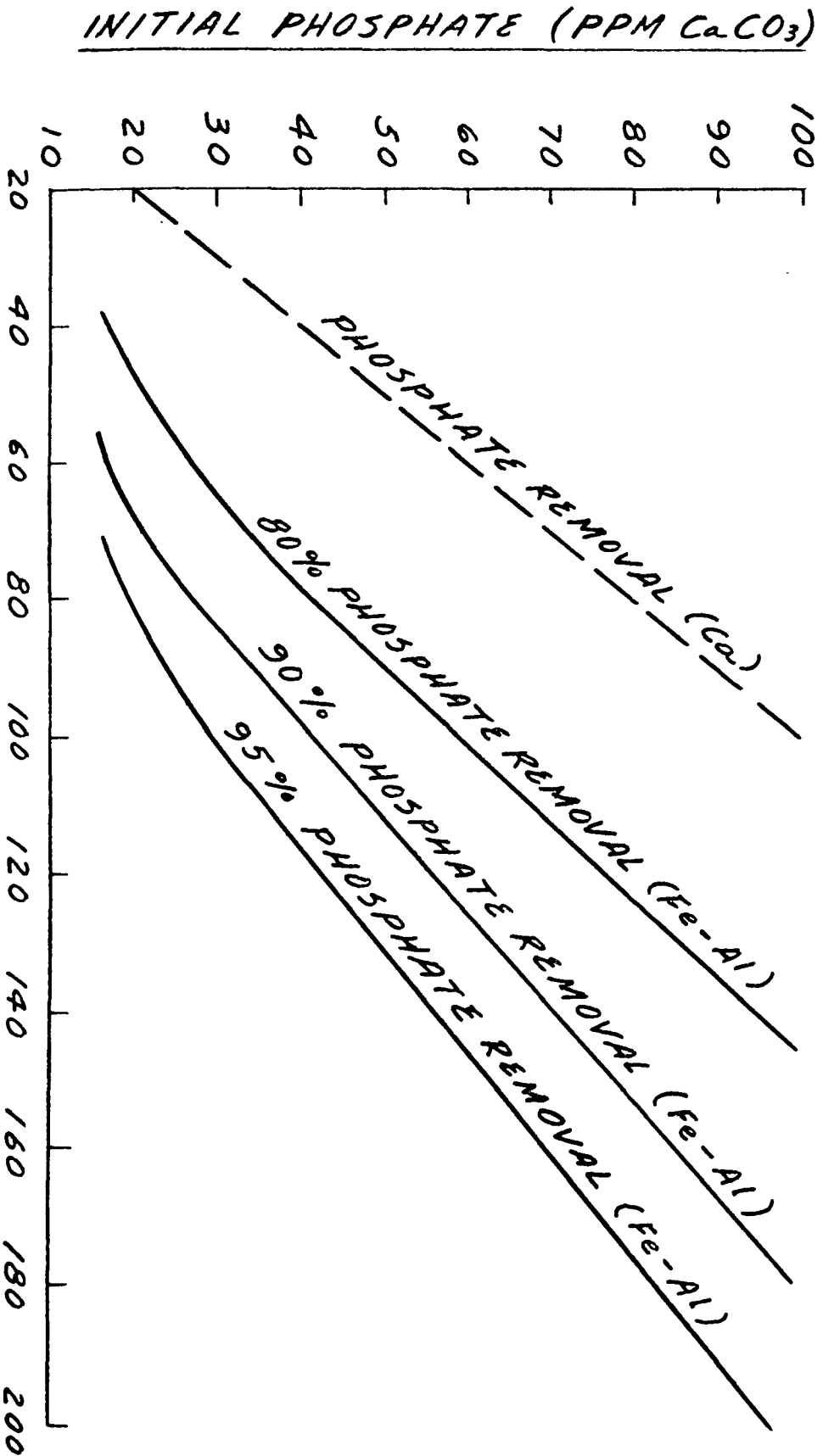
FIGURE 2

FIGURE 3

PROCESS DESIGN REQUIREMENTS

1. OPTIMUM TREATMENT CHEMICAL DOSAGE.

2. OPTIMUM ORDER OF ADDITION OF TREATMENT CHEMICALS.

3. RAPID, COMPLETE MIXING OF CHEMICALS, PREVIOUSLY FORMED PRECIPITATES AND INCOMING WATER.

4. GENTLE MIXING TO PROMOTE PARTICLE GROWTH.

5. ADEQUATE TIME FOR REACTIONS.

6. MINIMIZE SCALING OF EQUIPMENT SURFACES.

7. SEPARATION OF PRECIPITATES FROM WATER.

8 THICKEN AND REMOVE PRECIPITATES.

PHOSPHATE REMOVAL REACTIVATORS

CHEMICAL INTRODUCTION

CLARIFIED WATER

RAW WATER INFLUENT

SEDIMENTATION

SLUDGE REMOVAL

G

SLUDGE RECIRCULATION

C

H

TREATED WATER EFFLUENT

CLEAR WATER SEPARATION
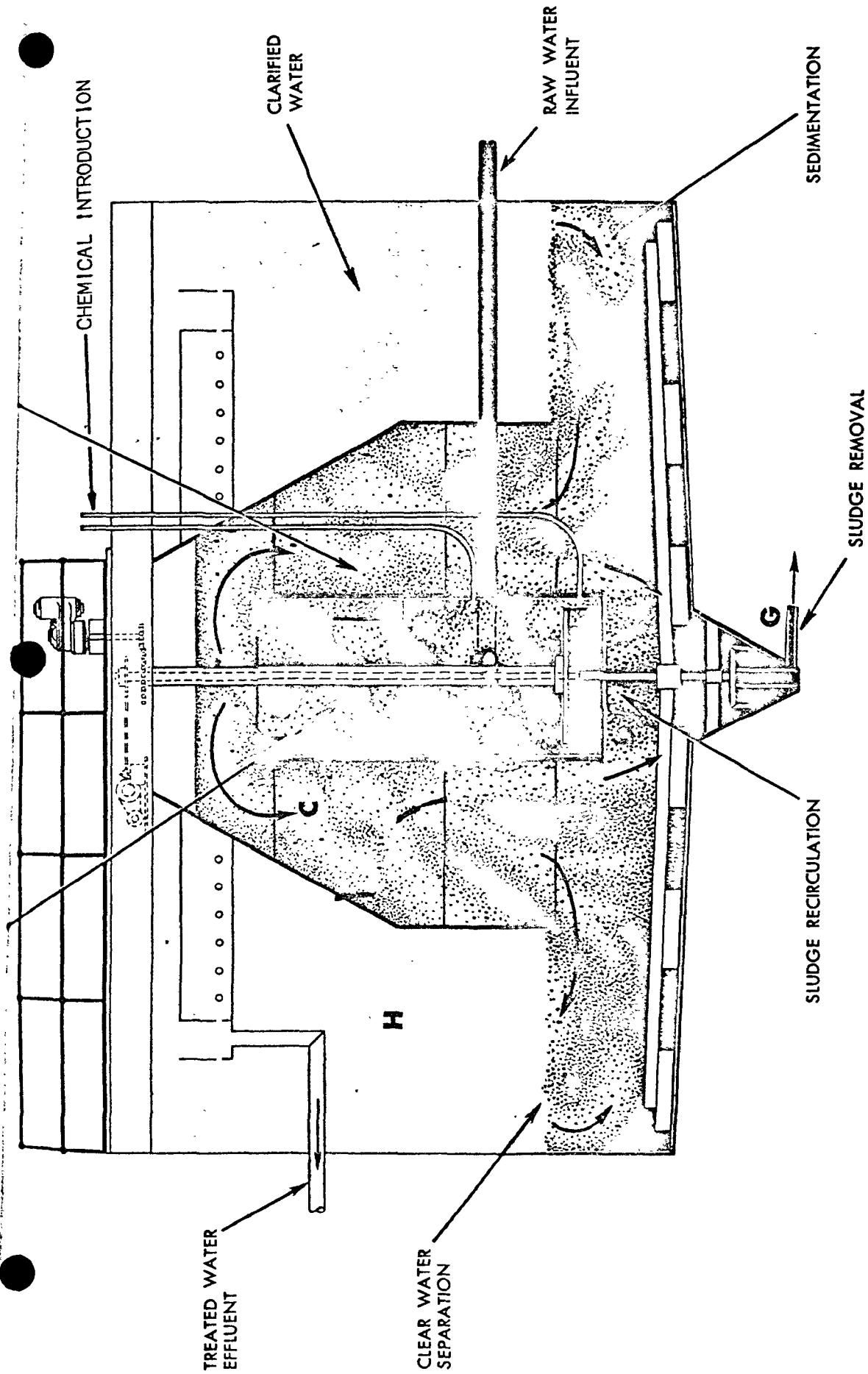
REACTIVATOR ® — *FUNCTIONAL DESIGN*

FIGURE 5

FIGURE 5

TYPICAL OPERATING RESULTS

	CITY WATER	SEWAGE PLANT EFFLUENT	REACTIVATOR EFFLUENT
CALCIUM-Ca^{++} (PPM $CaCO_3$)	110	140	140
MAGNESIUM-Mg^{++} "	130	160	80
SODIUM-Na^+ "	30	300	320
ALKALINITY (M.O.) "	230	300	320
SULFATE $SO_4^{=}$ "	20	30	—
CHLORIDE Cl^- "	20	300	—
PHOSPHATE $PO_4^{\equiv}$ "	0	40	1
SILICA (PPM SiO_2)	—	65	40
PH	—	7.9	10.5

FIGURE 7

POLYELECTROLYTE

LIME

ALUM

(ONE) 40'-0"∅
REACTIVATOR

(TWO) 10'-0"∅
DUAL MEDIA
AIR-SCOUR FILTERS

(FOUR) 8'-0"∅
GRANULAR ACTIVATED
CARBON ADSORBERS
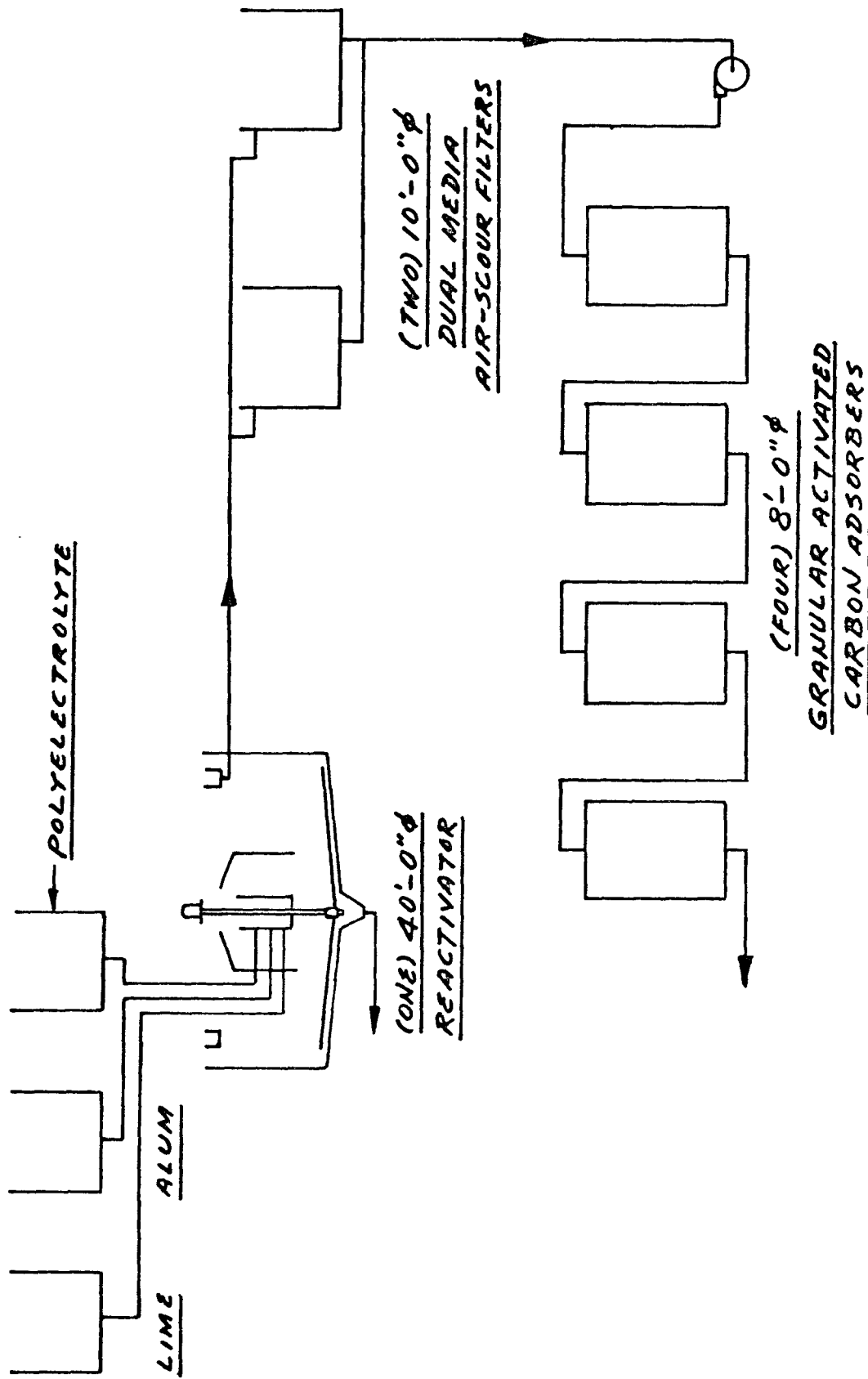
TERTIARY TREATMENT SYSTEM
NASSAU COUNTY

NASSAU COUNTY OPERATING RESULTS

(MAY 7, 1968)

		INFLUENT	REACTIVATOR EFFLUENT	FILTER EFFLUENT
CALCIUM (Ca^{++})	PPM $CaCO_3$	30	38	40
MAGNESIUM (Mg^{++})	"	37	26	26
ALKALINITY. M.O.	"	172	79	106
SULFATE ($SO_4^{=}$)	"	45	130	130
CHLORIDE (Cl^{-})	"	122	109	114
NITRATE (NO_3^{-})	"	2.2	1.8	—
PHOSPHATE POLY ($PO_4^{\equiv}$)	"	2.4	—	—
PHOSPHATE TOTAL ($PO_4^{\equiv}$)	"	34.4	6.8	0.4
IRON, Fe	PPM Fe	0.44	0.16	0.03
COD	PPM O_2	80	50	50
TURBIDITY (JTU)		50	6	0.2
PH		6.95	6.55	6.55

FIGURE 9

EQUIPMENT AND OPERATING COST

10 mgd PLANT

EQUIPMENT – INSTALLED COST

REACTIVATOR	$ 200,000
CHEMICAL FEEDS & HANDLING (LIME, ALUM, POLYELECTROLYTE)	45,000
DEWATERING EQUIPMENT	150,000
TOTAL	$395,000

OPERATING CHEMICAL COST

CHEMICAL	DOSAGE (LBS/M GALS)	COST C/LB	COST ¢/M GALS
LIME, (Ca O)	1.7	1.0	1.7
ALUM, $Al_2(SO_4)_3 \cdot 18H_2O$	0.25	3.0	0.75
POLYELECTROLYTE	0.003	100	0.3
TOTAL			2.75

POWER CONSUMPTION (100 % OF RATING)

REACTIVATOR AND CHEMICAL FEEDS – – – – – –	15 HP
DEWATERING EQUIPMENT – – – – – – – – – –	70 HP
TOTAL	85 HP
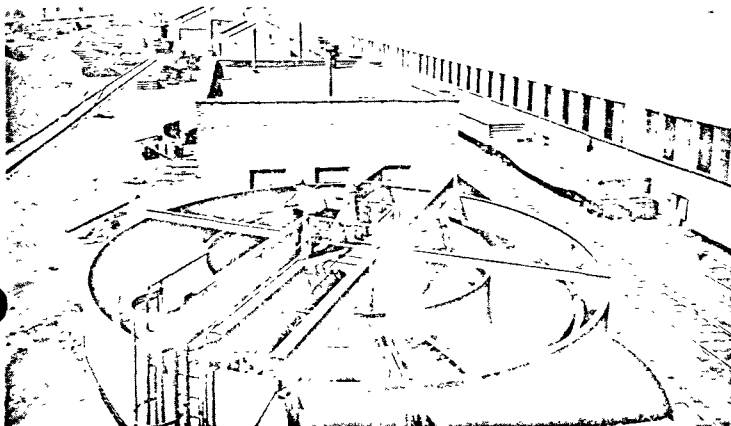
OIL REMOVAL In 1942 at the East Chicago, Indiana, plant of the Standard Forgings Corporation, a Graver clarifier 35' in diameter by 14' high was installed to remove oil and waste matter from a plant waste flow of 750,000 gpd.

SUSPENDED SOLIDS In 1944, at the Bakelite Corporation plant in Bound Brook, N. J., a Graver Filter System was installed to remove suspended solids from organic waste.
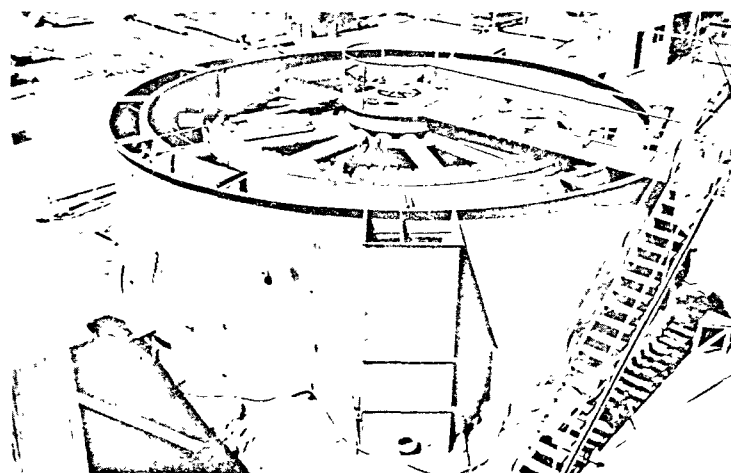
SEWAGE PLANT EFFLUENT In 1946, Socony Vacuum, Augusta, Kansas, installed a Graver clarifier to treat sewage plant effluent and other streams. Graver equipment was also put into operation for fluoride removal and the treatment of refinery wastes.

ELECTROPLATING WASTES In 1952, ion exchange was used for chromic acid recovery at Channel Master Corporation, Ellenville, N. Y.

For more than a quarter of a century, Graver has been providing systems and equipment to treat industrial wastes. The successful operation of equipment to remove suspended, colloidal and dissolved solids from industries as diverse as metal finishing, printing, paper making, fiber manufacturing, steel mills and oil refineries attests to the experience and knowledge acquired by Graver over the years.

Graver offers sound equipment design, authoritative process and equipment selection, efficient field engineering, highest quality fabrication and construction . . . assuring you and your consulting engineer of undivided responsibility for solving waste treatment problems. Process efficiency is assured through extensive laboratory and pilot plant research facilities.
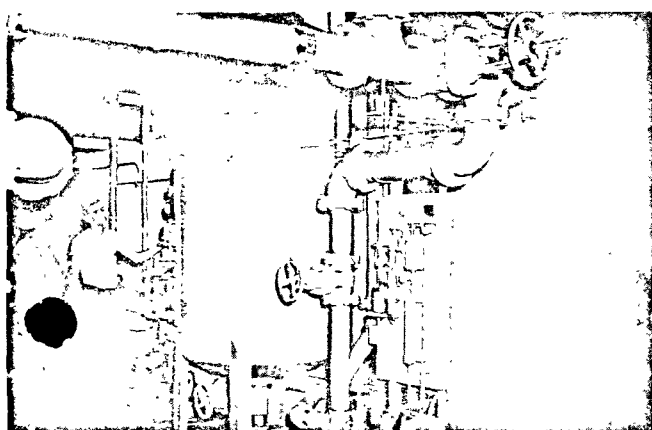
CYANIDE DESTRUCTION   In 1954, a continuous automatic cyanide plant went on stream at IBM's Endicott, N. Y., plant.
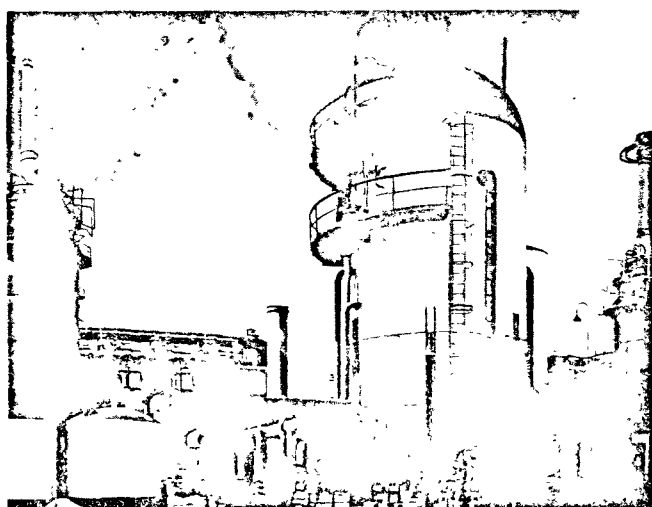
SEWAGE PLANT EFFLUENT RE-USE FOR BOILER FEEDWATER   In 1956. El Paso National Gas Products Co., Odessa, Texas, used sewage plant effluent, treated by clarification, filtration and ion exchange, for boiler feedwater.
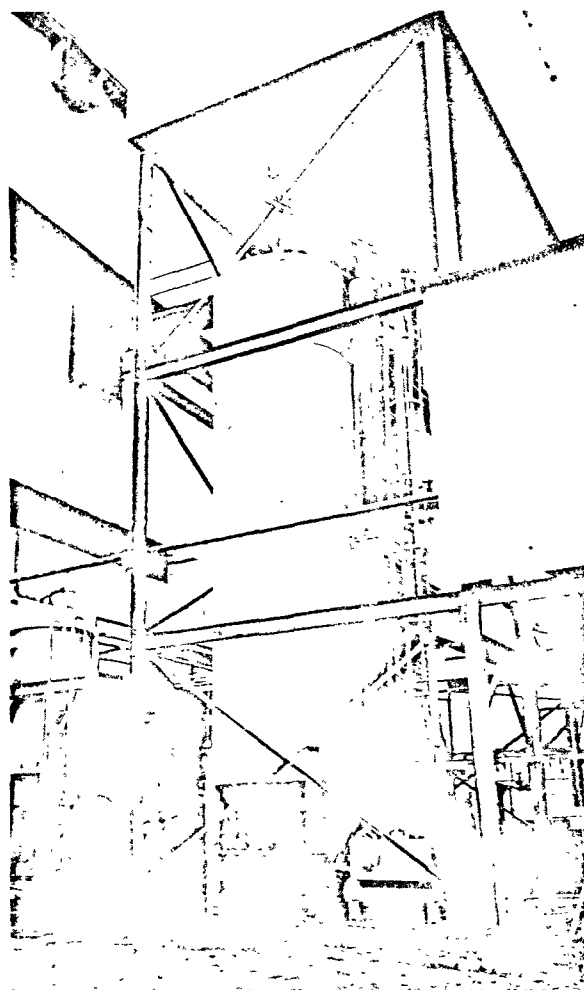
ORGANIC (POLYOL) REMOVAL   In 1964, Wyandotte Chemical Company, Washington, N. J., used continuous adsorption to treat polyols in solution and make the waste water suitable for discharge. A 4′ diameter, 20′ high Continuous Adsorption column, employing granular activated carbon, treats 70 gpm (100,000 gpd).

The Mono-Scour® filter is a high rate automatic filter particularly designed to operate with high suspended solids loadings. This is possible because the filter employs dual media and air scour. Experience has indicated that when conventional filter media are preceded by roughing filter media, much longer operating cycles and higher rate capabilities are obtained.

The Graver Mono-Scour filter offers two-stage filtration, i.e. a relatively coarse, low density material at the top of the filter bed, and a finer particle polishing layer of high density materials at the bottom of the filter bed. The Depth Filtration obtained with combination media permits high solids accumulations throughout the depth of the filter bed. Certain high solids accumulations within the filter bed may not be effectively removed by conventional backwash. Efficient air scour, in conjunction with backwash, is used in the Mono-Scour filter to insure positive bed cleaning.

Ordinarily, subfill-less underdrain systems are effective when the turbidity load applied to the filter is relatively light. Experience indicates that for heavy-duty multi-media filtration, conventional methods are frequently inadequate to prevent localized dirt accumulations in the lower portions of the bed. Graver Research has successfully developed the Roto-Scour System, a unique underdrain system able to meet the demanding requirements of heavy-duty high rate filtration. This system has its greatest application where the turbidity load to the filters is high, or where high filtration rates are employed.

One condition that impairs filter efficiency is the gradual build-up of dirt-impregnated media between strainers that remain untouched by ordinary subfill-less underdrain systems. The Roto-Scour Underdrain System provides great turbulence in the entire lower portion of the media bed, scrubbing it and eliminating "dead" areas. This scrubbing action is induced by the swirler plate that imparts a rotary sweeping action to the water and media during backwash. Complete bed cleaning is possible with the advanced Roto-Scour System.
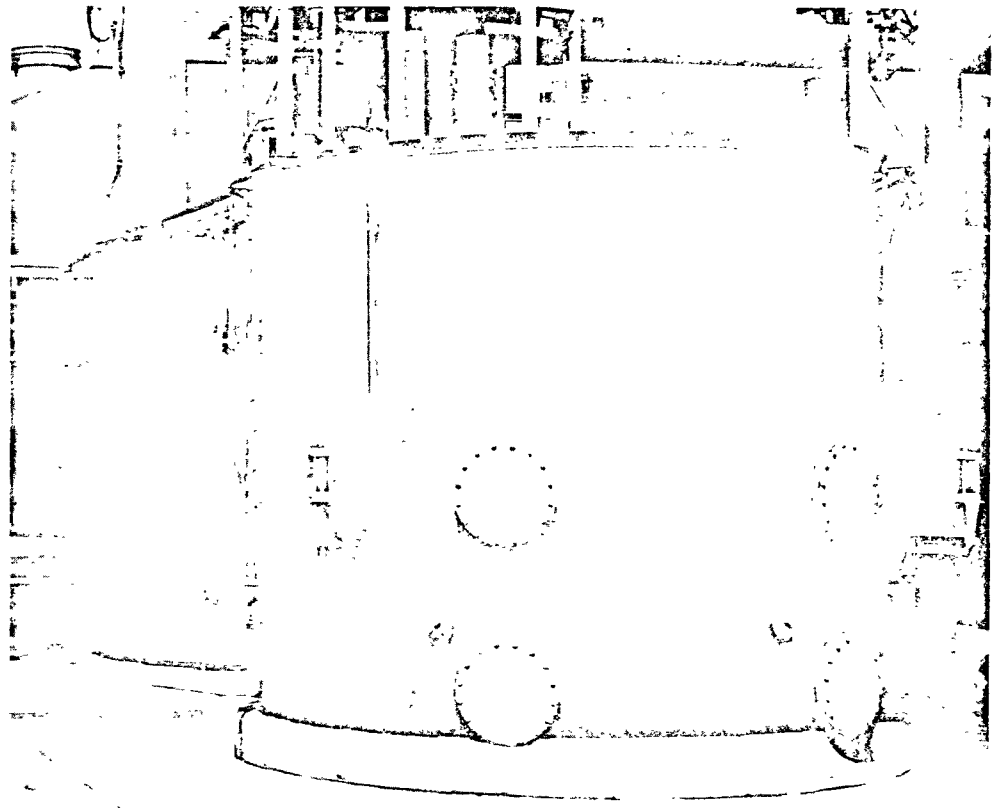
The Roto-Scour® Underdrain System is generally provided with an air scour system as a separate bed cleaning step. This air scour step is particularly necessary for heavy-duty service.

Many wastes previously handled by clarifiers, such as the effluents from steel mills, merchant mills, sewage plants (tertiary), etc., are now treated by filtration at considerable savings in operating costs, space and capital investment. (Write for Bulletin WC-133A for additional information).
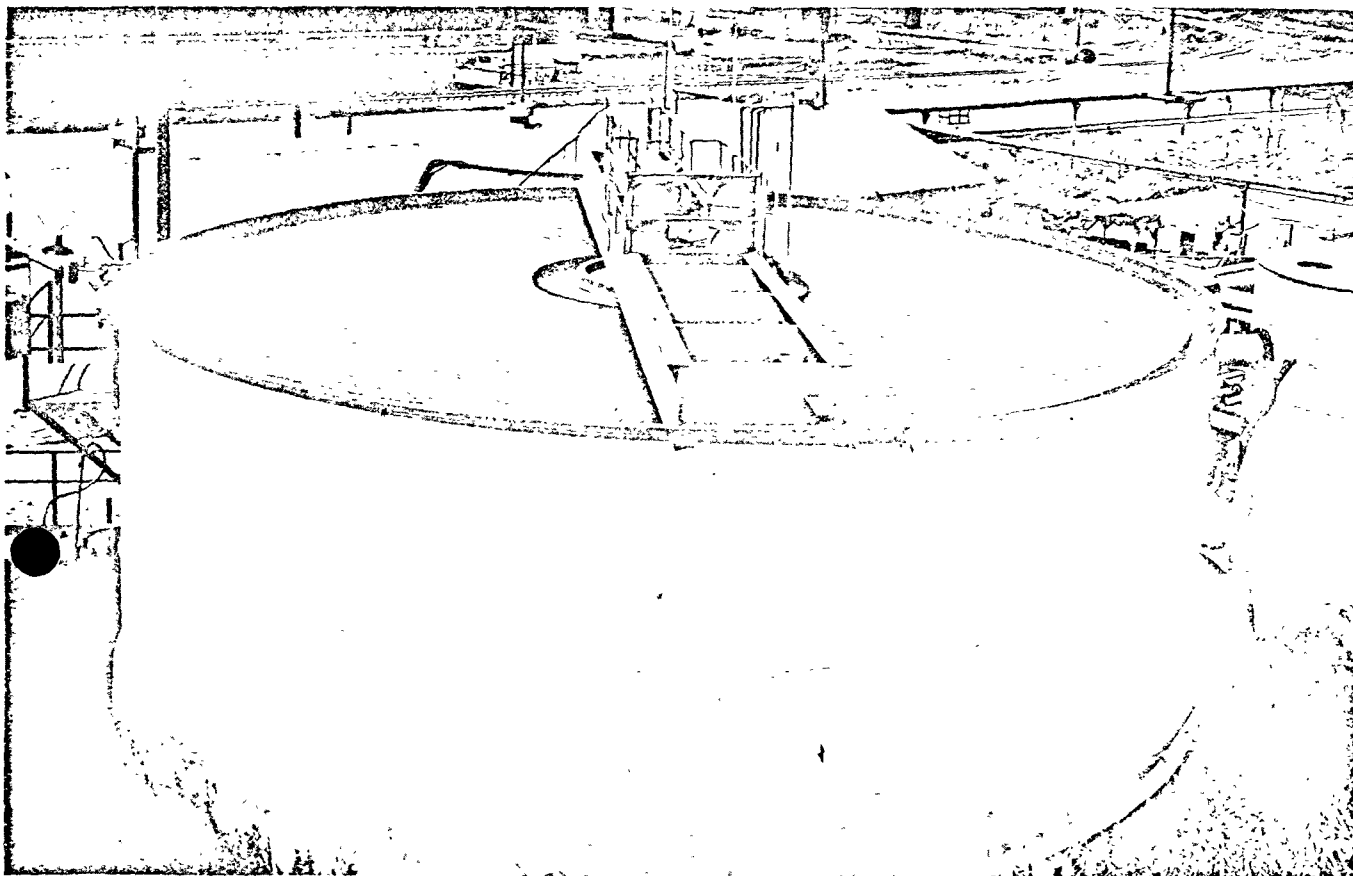
The Graver Reactivator* is a high-rate solids contact sludge recirculation clarifier ideally suited for chemical coagulation and clarification of waste water. It combines flash mixing, flocculation, clarification and sludge thickening into one operation.

When raw water enters the Reactivator, it is mixed with previously formed precipitates and treatment chemicals. The benefit of intimate contact with solids is obtained by the full retention time provided in the mixing zone, under the conical hood, so that by the time the water enters the outer settling zone, the bulk of the precipitated particles is large and dense.

The Graver Reactivator sludge removal system is especially advantageous when treating waste waters. Designs that use elevated open-lip concentrators, to which the suspended solids must be carried hydraulically, frequently fail because part of the sludge accumulates on the bottom and in time builds up so that the lower ports become completely plugged. When this occurs, it is necessary to shut down, and shovel or sluice out the accumulated sludge.

In the Graver Reactivator, the precipitated solids are moved mechanically by means of scrapers across the entire bottom of the unit to a central sump, thereby providing positive sludge removal. The Graver sludge removal system, including the slowly rotating scraper, central sludge sump, and automatic backflush and blowoff arrangements, is trouble-free in operation and easy to control. The ability of the Reactivator to recirculate, collect, thicken and remove sludge, makes this machine particularly applicable to sewage tertiary treatment, metal finishing, paper mills, oil refineries and steel mills. (Write for Bulletin WC-103D for additional information).
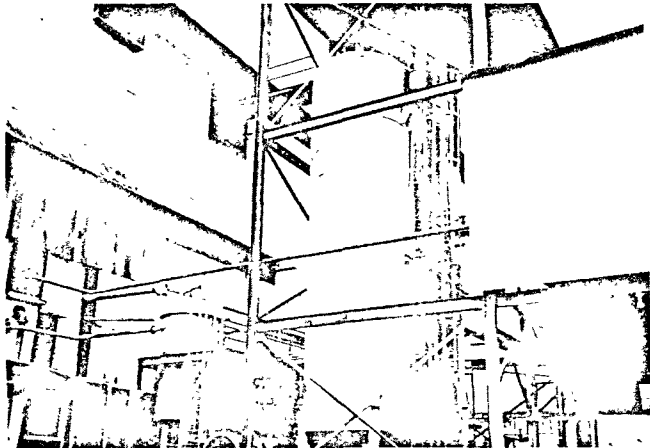
The Continuous Adsorption System is a fully auto-mated countercurrent system employing adsorbents in granular form. It is used for virtually complete re-moval of organics, particularly refractory materials.

Waste liquid is introduced continuously at the bot-tom of the C.A. unit and flows upward through the bed of activated carbon. A carbon retainer and upper collector allow regenerated adsorbent to be fed into the top of the unit, while treated water is discharged.

This operation permits the most regenerated ad-sorbent to be in contact with the waste stream just prior to its discharge. Acting as a "polishing" step, this last contact cleans the waste stream to a greater degree than is possible with fixed-bed adsorption systems.

The contaminant-saturated carbon is dewatered and then regenerated in a multiple hearth furnace. In this way, the carbon is re-used approximately 30 times before new carbon is required. Colored wastes can be cleared economically using activated carbon in a C.A. System. This is applicable to dye wastes and other clear but colored waste streams, as well as to tertiary treatment. In addition, there appears to be a definite place for the C.A. System following biological treatment, to assure that the final ef-fluent will meet the stringent requirements for dis-charge set by regulatory agencies.
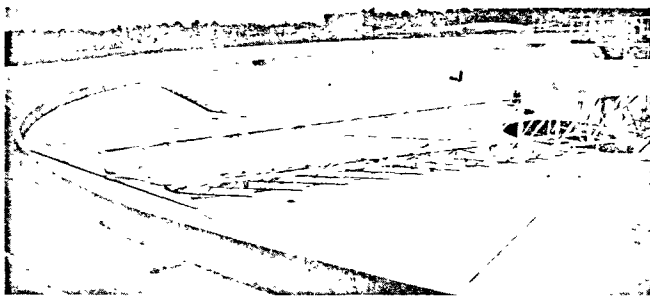
The Rota-Rake® provides for the gravity separation of suspended solids from liquids. It offers maximum separation, positive sludge removal, and complete overload protection. Available in a wide range of designs and sizes, the Rota - Rake is a simply de-signed, ruggedly built unit, economical to install, operate and maintain.

Rota-Rake installation consists of a round or square, conical bottom, steel or concrete tank with a quiescent flow feed arrangement and an overflow effluent system. Water is fed to a central inlet well and its velocity is reduced. Heavy solids are de-posited on the bottom of the tank. As the water flows radially across the unit, it settles out the other solids, the finest particles being deposited near the periphery where the velocity is lowest. The water then flows into collecting launders.

For positive sludge collection and discharge with a minimum amount of turbulence, a motor-driven, heavy-duty, box-truss scraper continuously moves settled solids to a central discharge sump in the bottom of the tank. A rotating paddle in the sump keeps the sludge moving and thickens it further.

The Rota-Rake is used as a primary clarifier for treating paper mill effluents. With the addition of appropriate skimming equipment, it is applicable for the removal of floating solids or free oil. Write for Bulletin WC-123A for additional information.
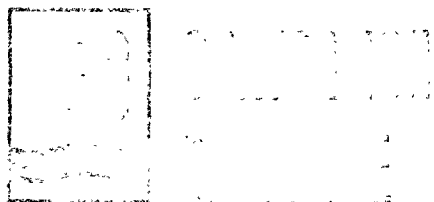
Effective removal of essentially all dissolved minerals from fluid industrial waste is accomplished with Graver Demineralizers. Demineralization, either fixed bed or by the Graver CI Process (continuous countercurrent ion exchange), is used to convert industrial waste water into the highest quality process water and boiler feed-water. As a by-product of this water purification, valuable materials are occasionally recovered. In some cases, the cost of the demineralization equipment will be more than offset by this material savings. Waste treatment applications include chromate and other plating metals recovery, concentration of rare metals and rare earths, and recovery of copper and zinc.

In addition to demineralizers, Graver also offers other ion exchange equipment including softeners, dealkalizers and disilicizers. A complete line of packaged demineralizers using the unique Partilok Strainer subfill-less underdrain system and the Monotrol® valve is also available. (Write for Bulletin WC-111B for additional information).

| USER	SERVICE	USER	SERVICE	USER	SERVICE
American Brass Co. Paramount, California	Metal Finishing (Chrome, Acid)	**General Motors Corporation New Departure Division** Bristol, Connecticut	Oily and Plating Waste	**Pine Castle Air Force Base** Pine Castle, Florida	Aircraft Washrack
American Cyanamid Fort Worth, Texas	Sodium Sulfate Recovery	**General Motors Corporation Euclid Division** Hudson, Ohio	Oily and Metal Finishing	**Pittsburgh Plate Glass Co.** Shelby, North Carolina	Process (Organic)
Armstrong Cork Co. Lancaster, Pennsylvania	Process Water (Cork Dust)	**General Motors Corporation Diesel Equipment Division** Grand Rapids, Michigan	Oily Waste	**Pyle National Co.** Aiken, South Carolina	Plating Wastes
Armstrong Cork Co. Lancaster, Pennsylvania	Paint	**General Motors Corporation Ternstedt Division** Warren, Michigan	Plating	**Rayonier, Inc.** Fernandina Beach, Florida	White Water
Avco Manufacturing Co. Stratford, Connecticut	Metal Finishing (Chrome, Cyanide, Acid, Alkali)	**Hawthorne Paper Co.** Kalamazoo, Michigan	White Water	**Rohm & Haas** Deer Park, Texas	Trickling Filter Effluent
Borg Warner Corporation Ingersoll Steel Corp. New Castle, Indiana	Steel Mill	**Hercules Inc.** Wilmington, Delaware	Acetic formic acid Waste-deepwell disposal	**Schrader Valve Co.** Dickson, Tennessee	Cyanide neutralization and Clarification System
Bridgeport Thermostat Division Robert Shaw-Fulton Control Co. Milford, Connecticut	Plating (Chrome, Cyanide, Acid, Alkali)	**Hercules Inc.** Wilmington, Delaware	Cotton Wash Water	**Schrader Valve Co.** Wake Forrest, North Carolina	Plating Wastes
Bristol Brass Co. Bristol, Connecticut	Metal Finishing (Chrome, Acid)	**Hercules Powder Co.** Hopewell, Virginia	Cotton Linters (Process Water)	**Shaw Air Force Base** Shaw Field, South Carolina	Aircraft Washrack
Burndy Engineering Co. Milford, Connecticut	Metal Finishing (Chrome, Cyanide, Acid, Alkali)	**Holland Color & Chemical Co.** Holland, Michigan	Dye	**Sikorsky Helicopter Division** Stratford, Connecticut	Metal Finishing
al Tex Oil Co. ermany	Refinery Wastes	**Homestead Air Force Base** Homestead, Florida	Aircraft Washrack	**Small Tube Products** Altoona, Pennsylvania	Chromic Acid
Carbide & Carbon Chemical Co. Torrance, California	Oily Waste	**Hunter Air Force Base** Savannah, Georgia	Aircraft Washrack	**Socony Mobil Oil Co.** Beaumont, Texas	Coking Waste
Carbide & Carbon Chemical Co. Whiting, Indiana	Oily Waste	**Inland Steel** E. Chicago, Ind.	Mill Waste	**Socony Vacuum Oil Co.** Augusta, Kansas	Refinery—Tertiary Treatment
Cessna Aircraft Co. Wichita, Kansas	Chromic Acid	**International Business Machines** Rochester, Minnesota	Metal Finishing (Chrome, Cyanide, Acid, Alkali)	**Standard Forgings Co.** Indiana Harbor, Indiana	Oily Waste
Channel Master Corp. Ellenville, New York	Chrome Plating	**Lincoln Air Force Base** Lincoln, Nebraska	Aircraft Washrack	**Stewart Air Force Base** Smyrna, Tennessee	Aircraft Washrack
Cross Keys Foundry Altoona, Pennsylvania	Chrome	**Lowe Paper Co.** Ridgefield, New Jersey	De-inking	**Stonebridge Paper Co.** Wilmington, Illinois	White Water
City of Dayton Dayton, Ohio	Tertiary Treatment	**McGuire Air Force Base** Wrightstown, New Jersey	Aircraft Washrack	**Superior Electric Co.** Bristol, Connecticut	Plating
Delta Airlines Atlanta, Georgia	Aircraft Washing and Plating	**Manchester Board & Paper Co.** Richmond, Virginia	White Water	**Texas Company** Anacortes, Washington	Refinery
Diamond Alkali Painesville, Ohio	Soda Ash Recovery	**The Martin Co.** Orlando, Florida	Metal Finishing (Chrome, Cyanide, Acid, Alkali)	**Triangle Conduit & Cable Co.** Landisville, New Jersey	White Water
Donaldson Air Force Base Donaldson, South Carolina	Aircraft Washrack	**Midwest Steel Co.** Portage, Indiana	Fluoride Removal	**United Shoe Machinery Co. S. O. & C. Division** Ansonia, Connecticut	Metal Finishing (Chrome Cyanide, Acid, Alk)
Doubleday & Co. Smithsburg, Maryland	Lithography	**Mobil Chemical Co.** Plainfield, New Jersey	Chemical Wastes	**U. S. Steel Corporation** Irvin Works Dravosburg, Pennsylvania	Oily Sludge
Downingtown Paper Co. Downingtown, Pennsylvania	White Water	**Nassau County-Bay Park Sewage Plant** Hempstead, New York	Tertiary Treatment	**Westinghouse Electric Co.** Pittsburgh, Pennsylvania	Copper M
Electric Autolite Co. Decatur, Alabama	Plating	**Pan American Petroleum** Edgewood, Texas	Refiner Waste	**Whippany Paperboard Co.** Durham Mill Riegelsville, Pennsylvania	White Water
El Paso National Gas Products Co. Odessa, Texas	Tertiary Treatment	**Phelps Dodge Co.** South Brunswick, New Jersey	Copper Mill	**Whippany Paperboard Co.** Stoneybrook Mill Whippany, New Jersey	White Water
Robert Gair Division Continental Can Co. Augusta, Georgia	White Water	**Piedmont Co.** Alta Vista, Virginia	Plating Wastes	**Whippany Paperboard Co.** Eden Mill Whippany, New Jersey	White Water
General Bronze Co. Garden City, Long Island, New York	Metal Finishing (Chrome Acid)			**Wyandotte Chemical Co.** Washington, New Jersey	Polymer Treatment
				Youngstown Sheet & Tube Youngstown, Ohio	Oily Waste

UNION TANK CAR COMPANY

WATER AND WASTEWATER TREATMENT LITERATURE

The following literature is available from the Smith & Loveless
division. Write to Robert J. Illidge, Advertising Manager,
Smith & Loveless division of Union Tank Car Company, Lenexa,
Kansas   66215.

Number	Subject
500-C	Pumps
601-A	Pumping Stations
605	"DupliFlo"
615-F	"Mon-O-Ject"
616-D	"Du-O-Ject"
630	"Vac-O-Ject"
2001-C	Check Valves
2010-C	"Shewer Tap"
1200-B	Factory-Built "Oxigest"
1220-C	Field-Erected "Oxigest"
1260-A	CY "Oxigest"
900-D	Comminutors
50-A	General Information
100-A	General Information
620	Largest Pump Station

Please note page 7 of the Graver Water brochure for literature
available on industrial waste treatment. This literature can
be obtained by writing to Martin Stern, Advertising Manager,
Graver Water Conditioning Co. division of Union Tank Car Company,
Highway 22, Union, New Jersey 07083.

Factory-Built Sewage Lift Stations and Sewage Treatment Plants
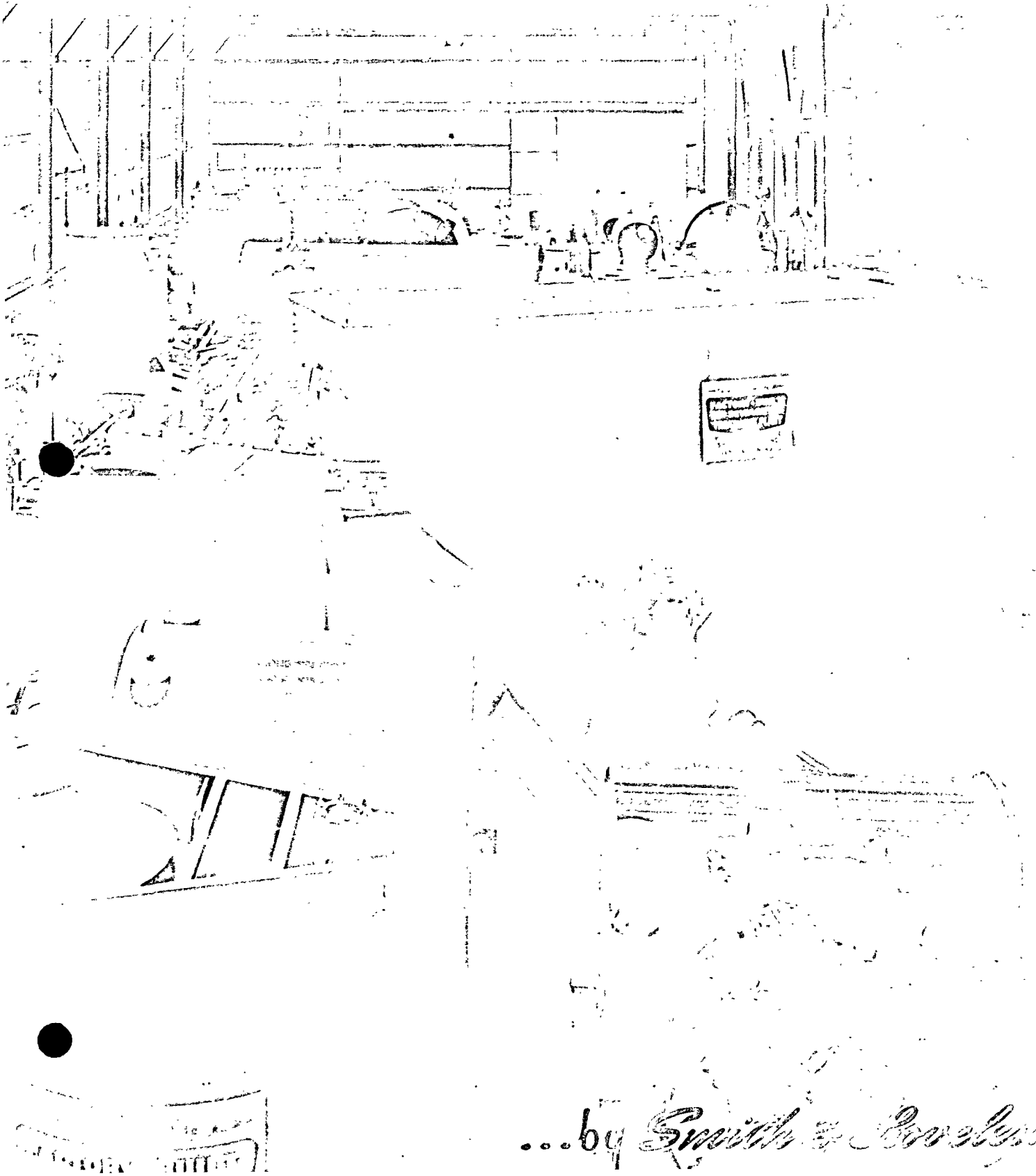
...by Smith & Loveless

HERE'S HOW THEY OPERATE: Beside the sewage pump station is a wet well or receiving manhole that receives the inflowing sewage. As the wet well level rises, the pipe ① of the air-bubbler system is submerged in the sewage, causing an increasing back pressure on the steady stream of air that is being ejected into the wet well by the compressor in the control panel ②. At a predetermined level, this back pressure in the system actuates a mercury pressure switch which energizes a magnetic starter. One of the pumps ③ goes on and the sewage flows out of the wet well, into the station, through the inlet pipe ④, through a gate valve ⑤, into the pump ⑥, out through a check valve ⑦, the discharge gate valve ⑧ and out the discharge force main ⑨.

The sensitive mercury pressure switches for each of the sewage pumps are set at different levels so when necessary all pumps will operate to handle maximum flows. An automatic alternator is provided to change the sequence of operation of the pumps on the completion of each pumping cycle. This allows even distribution of the work load on the pumps. It, too, provides for all pumps to operate simultaneously should the level of the wet well continue to rise above the starting level for the low-level pump cycle.

The system is automatic, efficient and dependable.

Complete specifications, job recommendations, installation instructions, detailed drawings, pump-selection charts, operating instructions and a representative list of installations can be found in the 100-page data manual on Smith & Loveless sewage lift stations.

SEE BULLETIN 601

16 mm. color-sound motion picture on factory-built sewage lift stations by Smith & Loveless is available through sales representatives in principal cities.

Smith & Loveless Factory-Built Sewage Pumping Stations are built to provide a lifetime of dependable service—available for capacities from 100 to 4,500 GPM per pump with two, three and four pumps per station with a wide variety of control systems including variable speed controls and flow-matching systems, ozone generators, elevators, and alarm systems. Even larger capacity stations are built to order.

The Smith & Loveless sewage pump station is a complete factory-built unit that includes all equipment ready to operate. Every item of mechanical and electrical equipment has been factory-tested by actual operation under simulated design conditions to assure proper operation in the field.
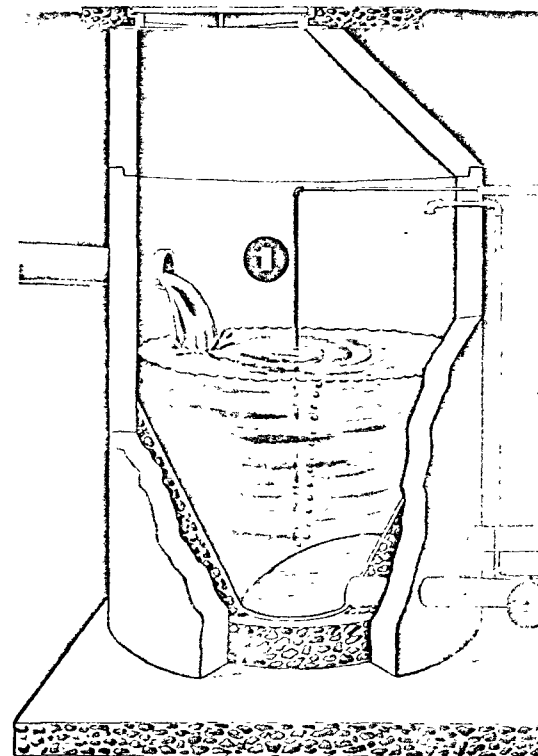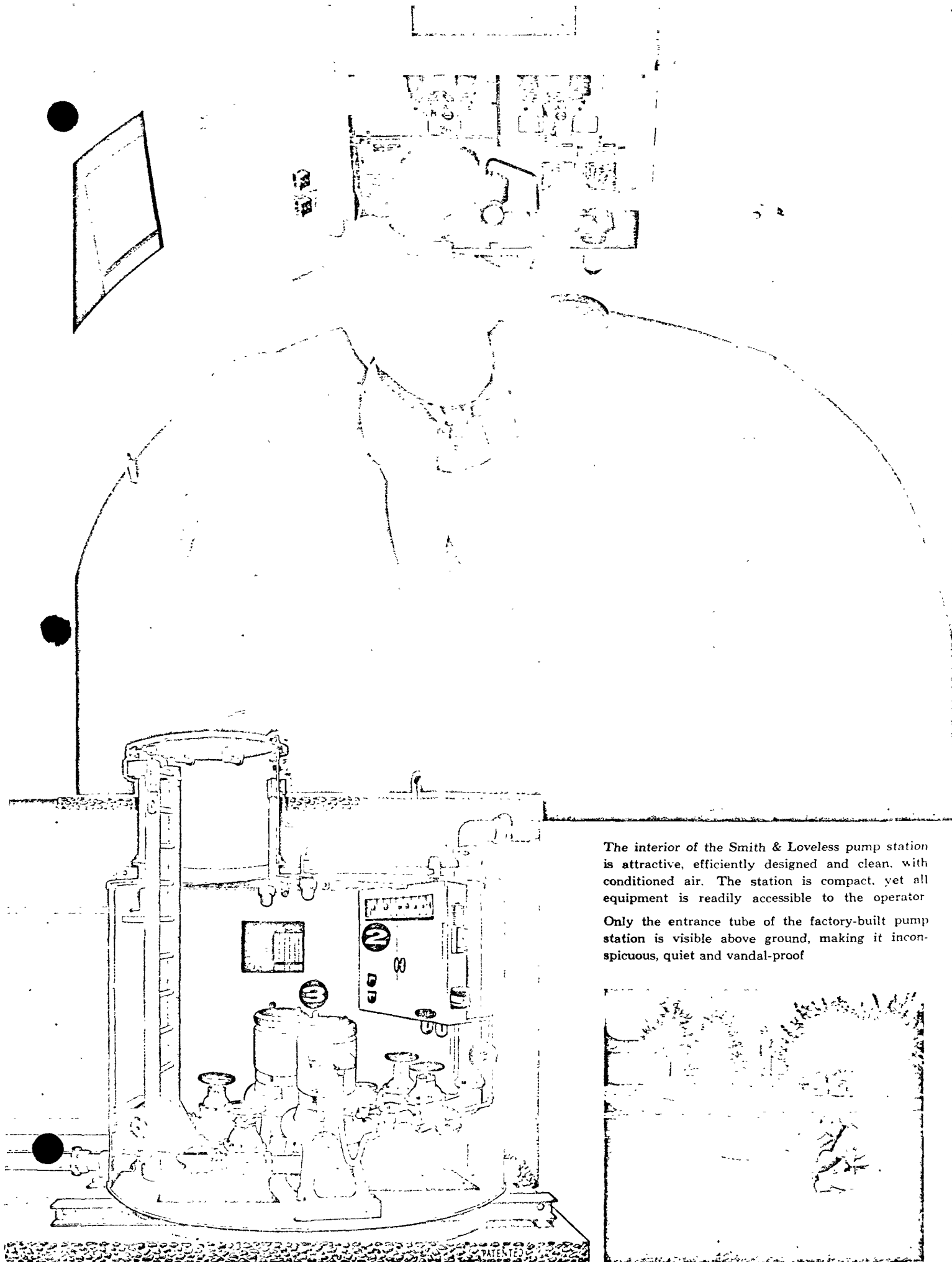
The interior of the Smith & Loveless pump station is attractive, efficiently designed and clean, with conditioned air. The station is compact, yet all equipment is readily accessible to the operator

Only the entrance tube of the factory-built pump station is visible above ground, making it inconspicuous, quiet and vandal-proof
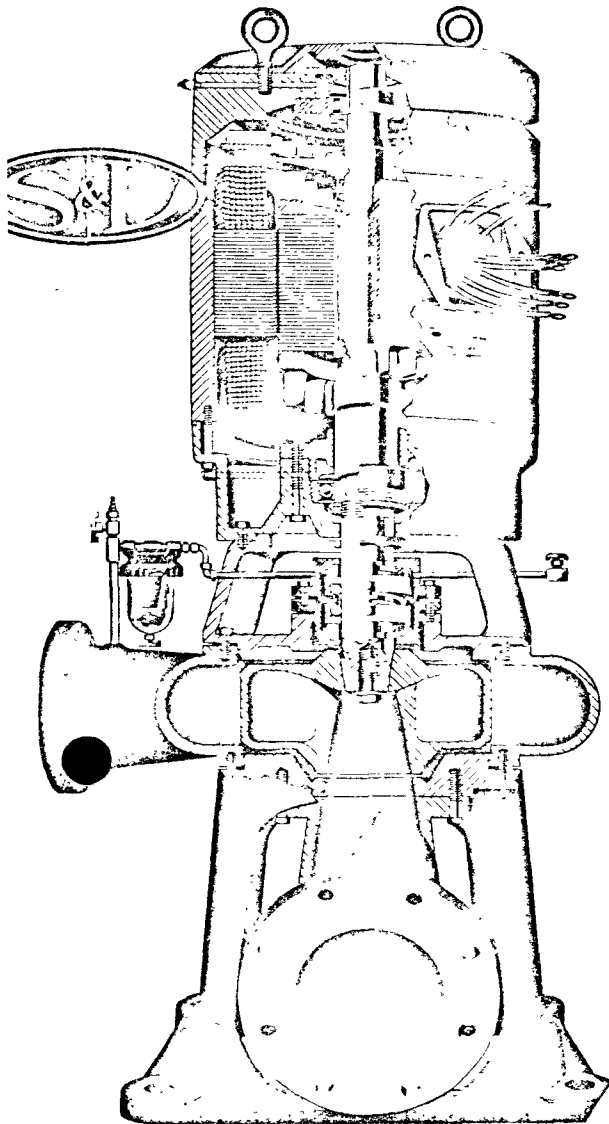
Featuring... the Smith & Loveless Vertical Close-Coupled "Non-Clog" Sewage Pump — the heart of the sewage pump station. Its design and construction reflect years of experience with over 2,500 varied pump installations by the world's largest manufacturer of factory-built sewage lift stations.

SEE BULLETIN 500-B

The double mechanical seal system in the Smith & Loveless "Non-Clog" Sewage Pump provides a "dead-tight" long-wearing seal that prevents sewage, noxious and dangerous gases from entering the pump room. The mechanical seal eliminates expensive replacement of worn-out shaft sleeves, and the leakage and maintenance problems of old-fashioned pump packing. When inevitable wear makes it necessary to replace the dead-tight seal, it can be done by one man in less than 30 minutes — a far cry from packing or seal maintenance on other pumps on the market.

Designed for easy maintenance, the Smith & Loveless "Non-Clog" Sewage Pump is built to close tolerances with heavy, rugged construction.

Motor Adaptor
Motor-Pump Shaft
Stationary Carbon
Rotating Ceramic
Lubrication
Stainless Steel Spring
Bronze Seal Housing
Tapered Shaft
Impeller Cap Screw
Keyed Shaft Fit
Impeller
Volute
Filter
Heavy-Duty Pump Base
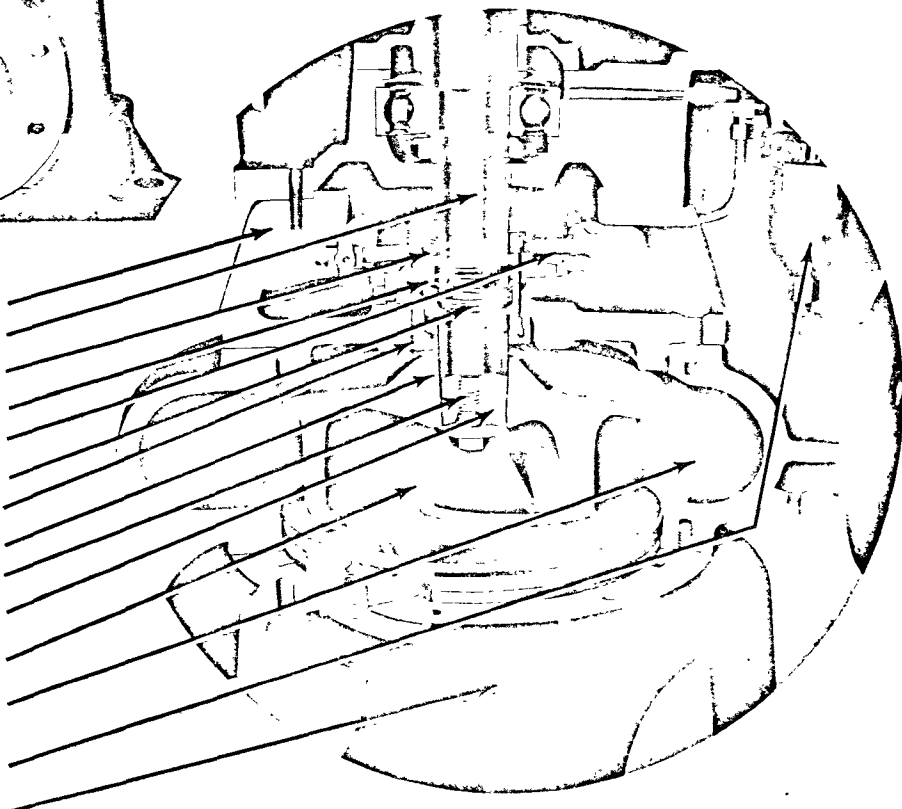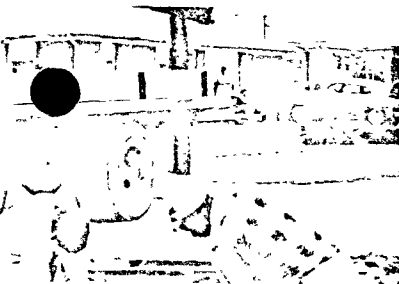
Smith & Loveless sewage pump stations are built "with the maintenance man in mind." Cover rungs extend the ladder for his safe entry, and when the cover is opened, the lights and ventilating blower turn on, automatically. The station comes complete with understandable maintenance and operating instructions, full color-coded wiring in the control panel, easy-to-reach starter and breaker switches — all fully identified for the operator's convenience and safety — a dependable, patented ventilating system and humidity control, plus ample "elbow room" so the maintenance man can work more efficiently. Every detail is designed for easier, faster maintenance and safer, more dependable operation.

Installed in Less Than a Day

Through proper production timing and coordination with the installing contractor, the Smith & Loveless sewage pump station can be scheduled for delivery at the installation site on time, when needed by the contractor, with all mechanical equipment ready to operate. The pump station is delivered to the job site on special-made trucks or by rail in the case of larger units. Complete installation and operating instructions are delivered with the pump station.
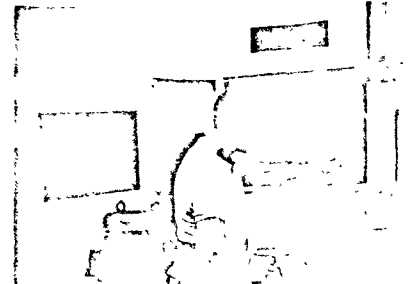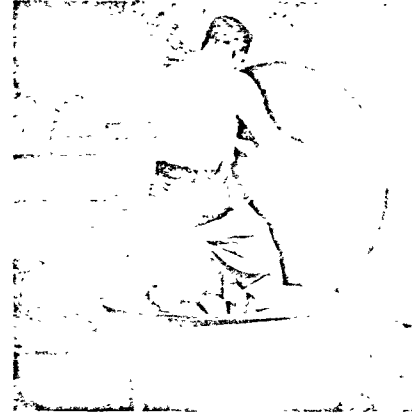
The Smith & Loveless "Mon-O-Ject" simple pneumatic ejector lift station is a cylindrical steel chamber with three individual compartments. The top compartment is easily accessible from ground level and houses the ejector controls and compressor. The middle section is a combination air-storage tank and chamber for the valves and manifold. The bottom compartment is the sewage receiver.

The pneumatic ejector lift station is recommended where the rated capacity is less than 100 GPM. A four-inch centrifugal pump, designed for three-inch solids, is the smallest which will operate with reasonable freedom from clogging on raw sanitary sewage. Such a pump is not available with a rated capacity less than 100 GPM. Therefore, to lift smaller flows, the pneumatic ejector is the only satisfactory answer. Only the larger sewage pumps can match the pneumatic ejector in its ability to handle large sewage solids.

SEE BULLETIN 615

TYPICAL INSTALLATION of a Smith & Loveless "Mon-O-Ject" sewage lift station serving a school in the suburbs of a growing community.
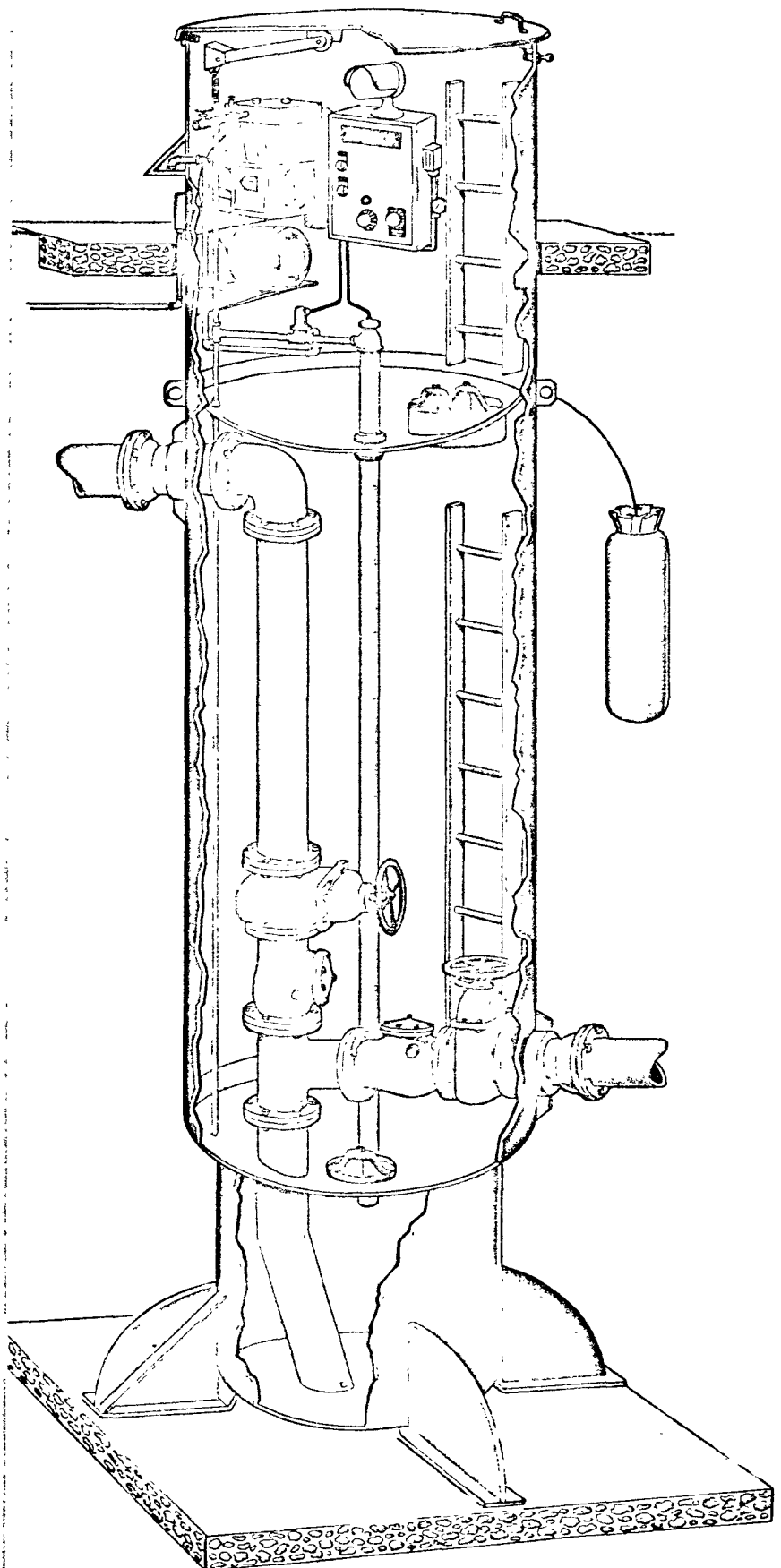
IDEAL FOR SMALL SUBDIVISIONS
MOTELS • SCHOOLS • FACTORIES

The Smith & Loveless "Du-O-Ject" duplex pneumatic ejector lift station is the equivalent to two "Mon-O-Ject" stations, combined into the same structure with dual piping and receivers. The "Du-O-Ject" provides the added dependability of complete stand-by equipment throughout and extra capacity for peak loads.

The "Du-O-Ject" is a cylindrical steel chamber with three individual compartments, housing the controls, the valves and the sewage receivers. The top compartment is easily accessible from ground level and houses the ejector controls and compressors. The middle section is a combination air-storage tank and chamber for the dual piping, valves and common discharge piping into the force main. The bottom section is divided into two separate compartments which are the sewage receivers.

SEE BULLETIN 616

Both Equipped with Exclusive
"No-Fail" Electrode System

An exclusive "No-Fail" Electrode System (Patents Pending)—composed of heavy-duty rectifiers and capacitors, ultra-sensitive DC relays, high-accuracy timing units and a uniquely designed, hollow electrode—prevents grounding and insulation failures that put conventional ejectors out of operation. The "No-Fail" system makes these units the most trouble-free sewage ejectors on the market today.

Complete specifications and information on the "No-Fail" Electrode System can be obtained by writing for the 100-page data manual on Smith & Loveless sewage lift stations.
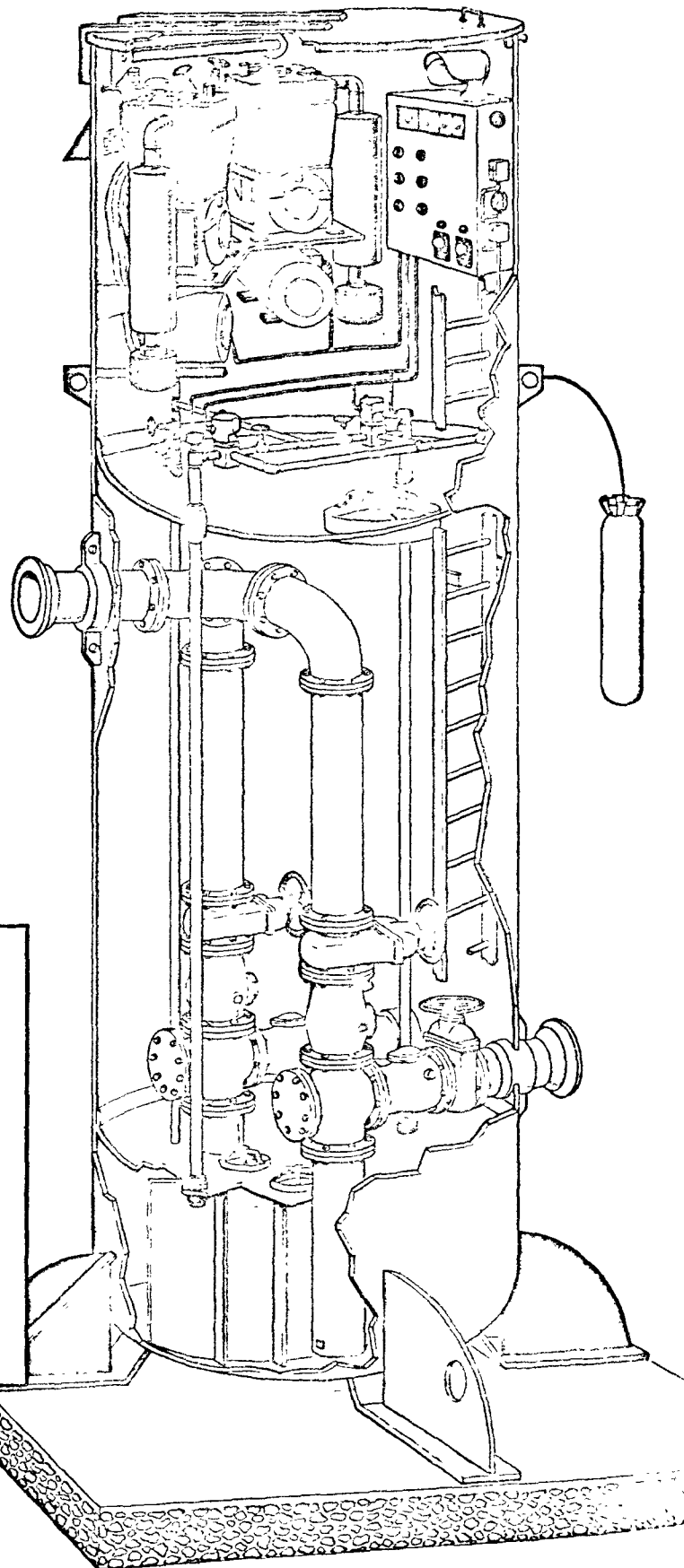
IDEAL FOR SMALL SUBDIVISIONS
MOTELS ● SCHOOLS ● FACTORIES

The Smith & Loveless "Way-O-Matic" pneumatic ejector lift station is constructed similar to the pump station but with ejector pots (receivers) replacing the pumps. A superior, patented control system "weighs" the sewage to provide dependable, trouble-free operation with minimum maintenance.

SEE BULLETIN 610

Sewage enters the station through the influent pipe ①, the inlet gate valve and check valve, into the sewage receiver ②. Air displaced by the sewage is vented out of doors through the three-way air valve ④ during the filling cycle.

As the receiver fills, the weight of the sewage causes the receiver to rotate slightly (see inset) around the pivot point. The unit is constructed to permit a maximum movement of about one-sixteenth inch.

When sewage reaches the high-water level, a high-level microswitch ③ makes an electrical contact which actuates a relay to the three-way air valve ④ closing off the vent line and opening the connection to the compressed air supply. The compressed air forces the sewage out the discharge manifold through the discharge check valve, gate valve and into the force main. ⑤.

As the sewage level falls, the receiver pivots back into the empty position, breaking the electrical contact and de-energizing the three-way air valve which shuts off the air supply and reconnects the vent line.

SIGNED FOR LOWER CAPACITIES
TO 200 GALLONS PER MINUTE
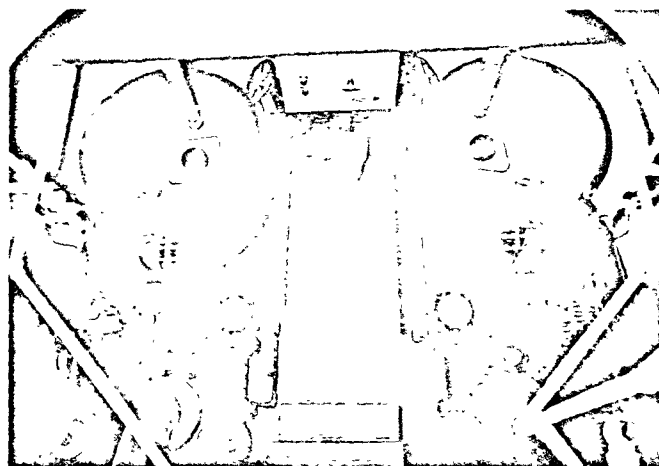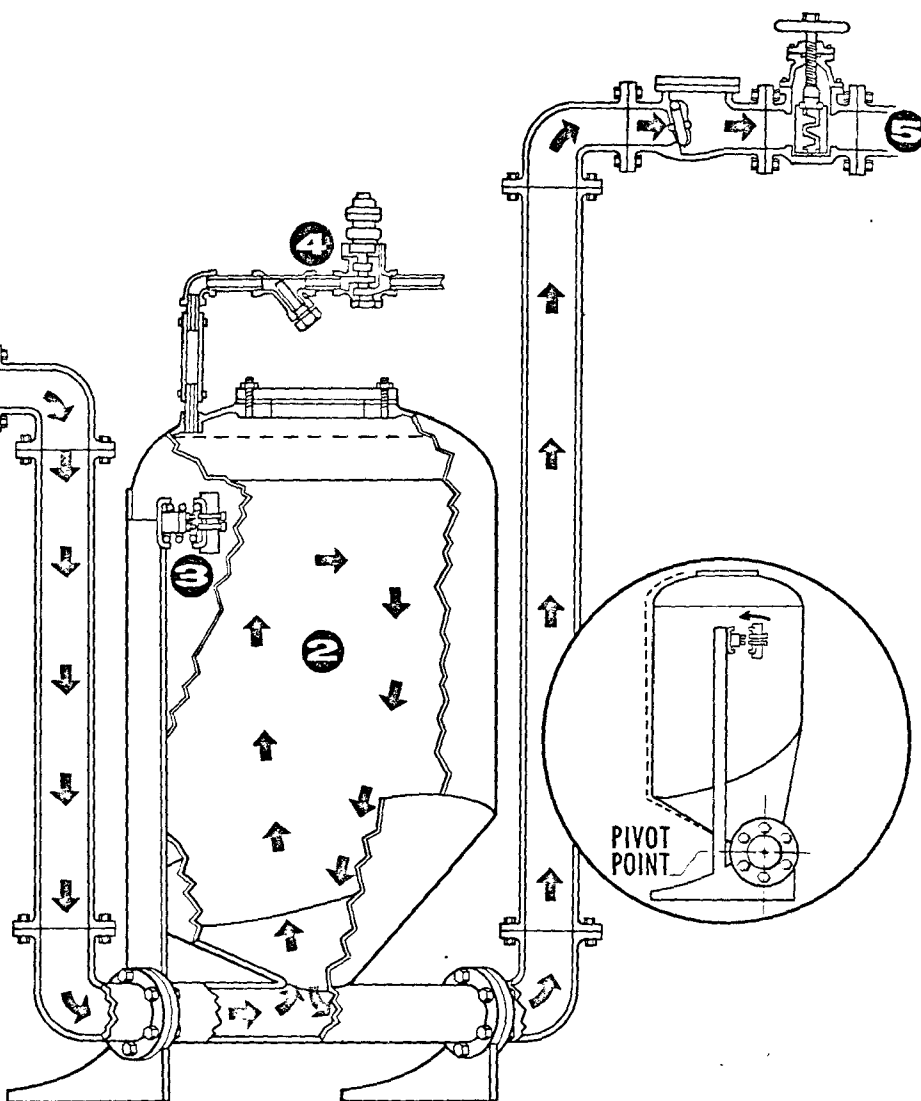
This interior view of the Smith & Loveless "Way-O-Matic" shows the arrangement of the ejector pots, control panel and compressors.

"VAC-O-JECT"
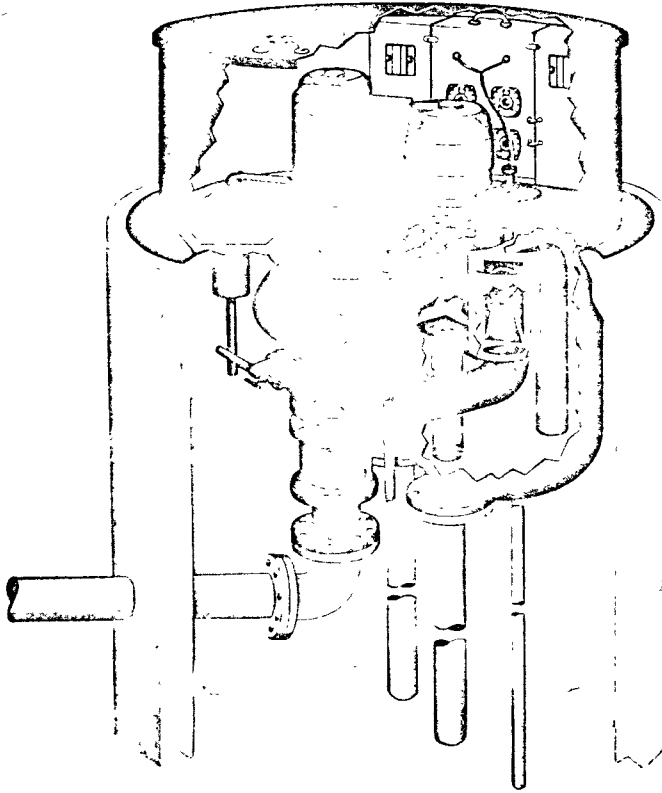
The Smith & Loveless "Vac-O-Ject" pneumatic ejector provides dependable two-stage lifting action, incorporating the principle of vacuum intake and air-pressure ejection. It is designed for installation on a standard four-foot wet well or receiving manhole.

As sewage rises in the wet well, the control system starts a reversible air pump which evacuates air from the receiver. Atmospheric pressure forces sewage up to the receiver. When filled, the air pump stops, reverses and ejects the sewage.

The "Vac-O-Ject" is specifically designed for lower capacities requiring an economical installation. It has the plus value of duplex dependability.

SEE BULLETIN 630

"PRES-O-JECT"
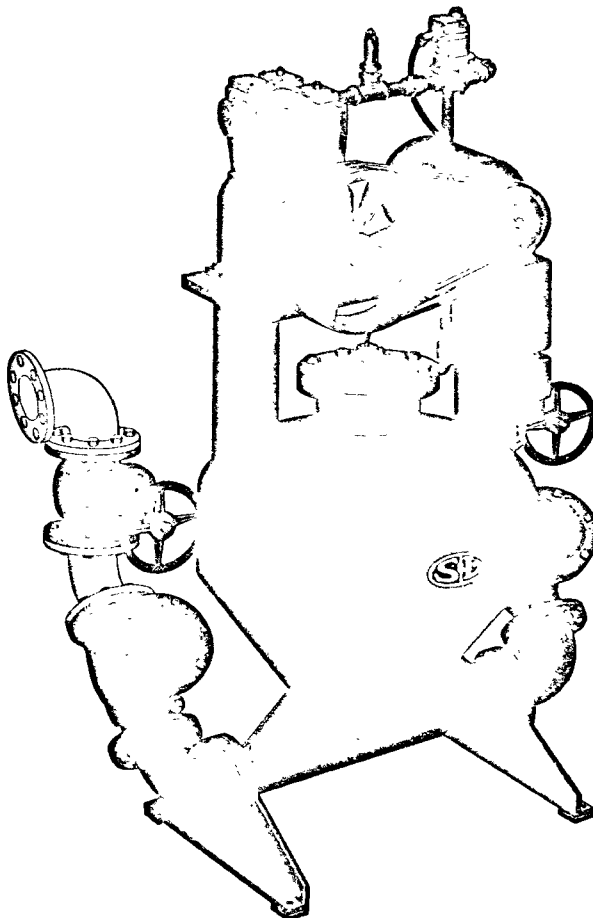
The Smith & Loveless "Pres-O-Ject" is a compact, factory-built pneumatic ejector. It can be specified with the conventional electrode system or the exclusive Smith & Loveless "No-Fail" Electrode System.

Sewage flows by gravity into the sewage receiver. When filled, an electrode actuates a Solenoid valve which closes off the vent line and starts the air compressor which forces air into the sewage receiver ejecting the sewage.

Designed for capacities up to 200 G.P.M., the "Pres-O-Ject" can be installed in simplex or duplex arrangements. Also available for stored-air applications and remote installation of compressors.

SEE BULLETIN 710

FACTORY-BUILT SEWAGE TREATMENT PLANTS

Smith & Loveless

"OXIGEST"®

WITH AUTOMATIC SURFACE SKIMMING*

Factory-built "Oxigest" sewage treatment plants are designed specifically for small subdivisions, mobile home courts, motels, apartments, shopping centers, resorts, hospitals, schools, factories and other small developments in outlying areas not served by municipal sewerage facilities.

The "Oxigest" with Automatic Surface Skimmer* is a major break-through in the maintenance and operation of small sewage treatment plants. The non-mechanical surface skimmer automatically removes floating material, via an exclusive Smith & Loveless hydraulic system, from the surface of the settling basin compartment. The recirculation effect eliminates operation problems, reduces maintenance.

Each diffuser assembly is readily removable (see left)—just one of the many advanced engineering features on the "Oxigest."

Available in a variety of standard sizes, "Oxigest" units can be installed in parallel, as needed, to keep pace with a growing sewage load from an expanding development. For complete engineering data on this "Aerobic Digestion" treatment plant, write for the 100-page data manual on Smith & Loveless sewage treatment plants.

Efficient, Dependable, Factory-Built

A 16mm, color-sound motion picture on Smith & Loveless products is available through sales representatives in principal cities.
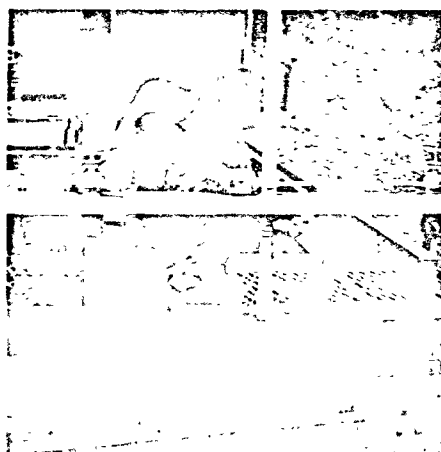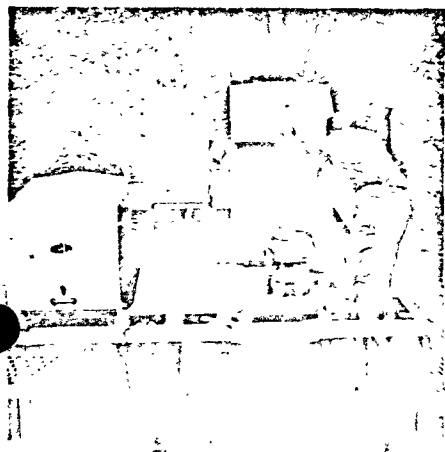
*Patents Pending   SEE BULLETIN 1200

Mike Microbe
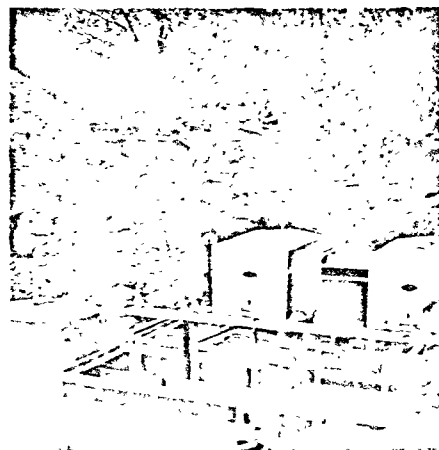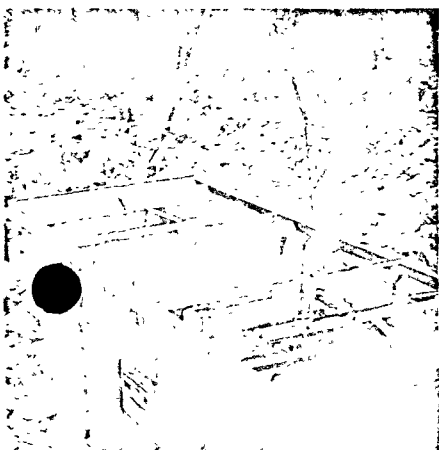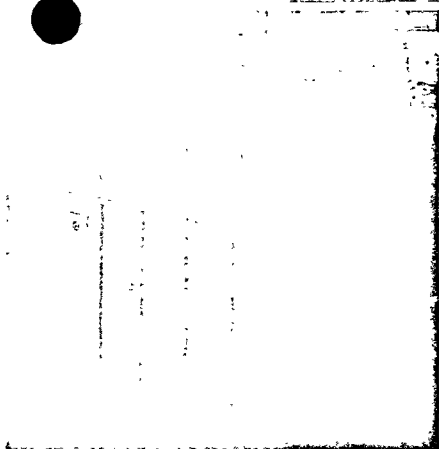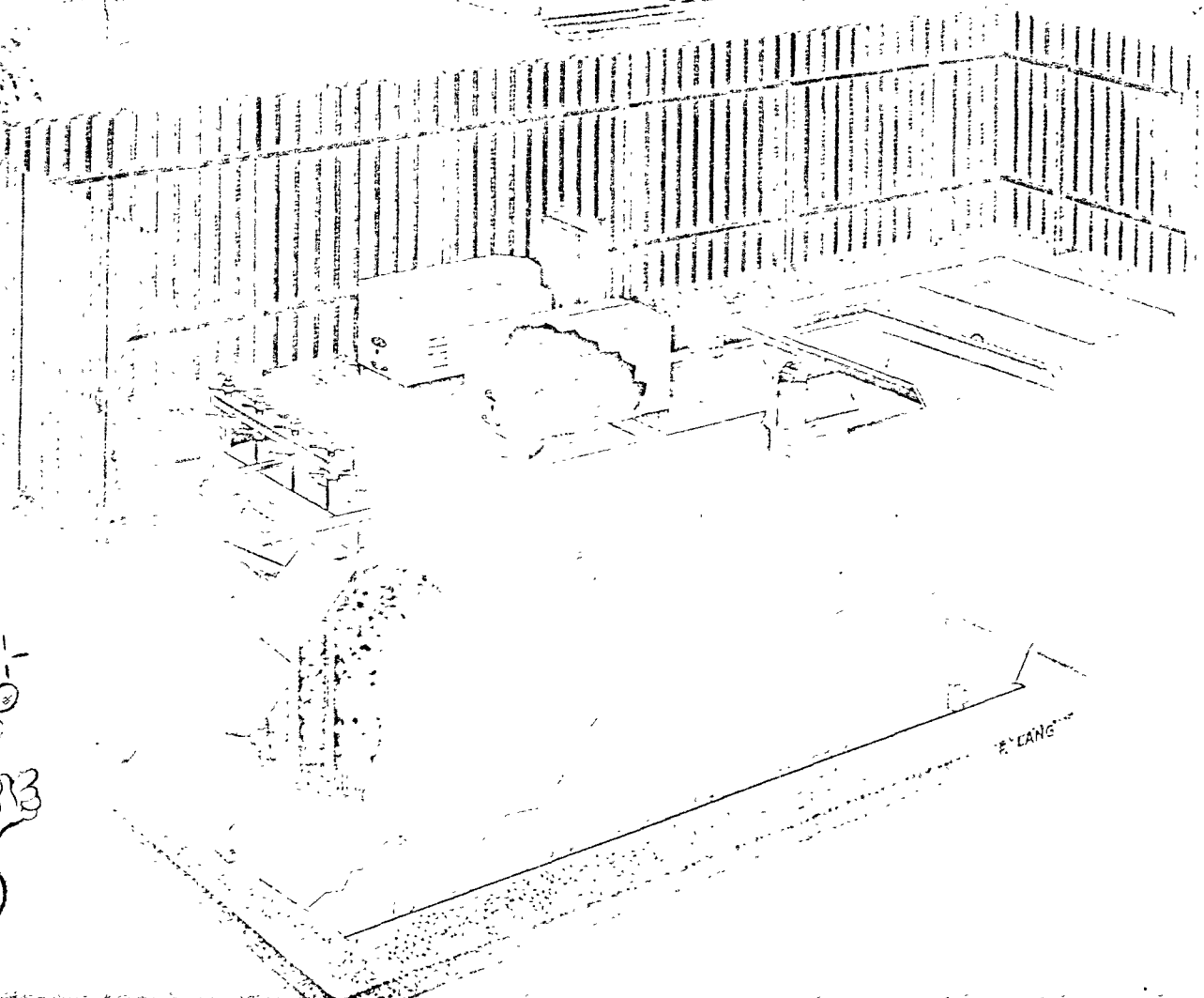
Dependable Sewage Treatment FOR MOTELS, MOBILE HOME PARKS, SMALL SUBDIVISIONS, FACTORIES, SCHOOLS, MILITARY BASES
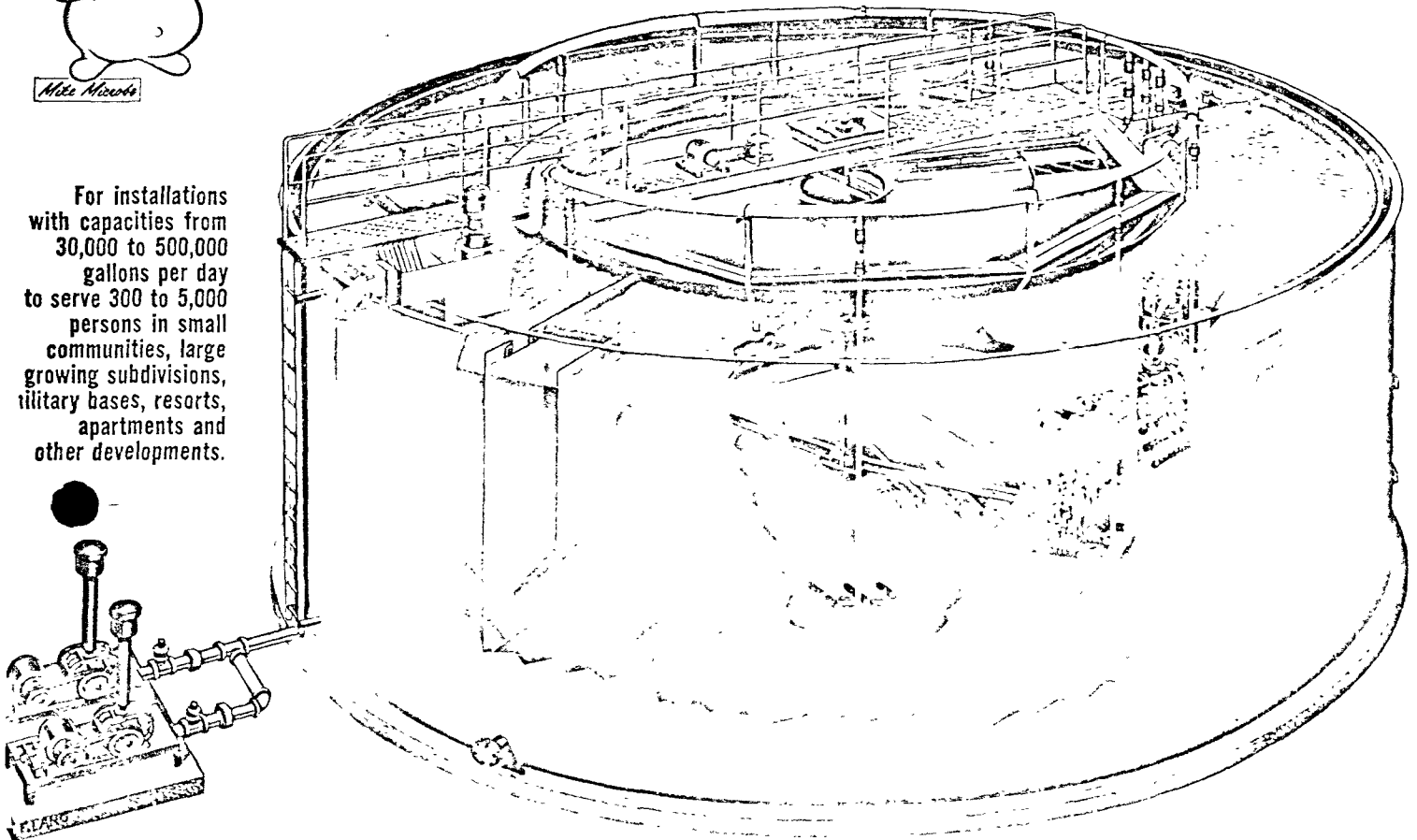
FIELD-ERECTED SEWAGE TREATMENT PLANT

Smith & Loveless
"OXIGEST"®
Designed for Larger Capacities

For installations with capacities from 30,000 to 500,000 gallons per day to serve 300 to 5,000 persons in small communities, large growing subdivisions, military bases, resorts, apartments and other developments.

Large field-erected "Oxigest" units provide dependable sewage treatment with minimum maintenance. The plants are factory-fabricated to be field-erected at the job site. The purchaser receives a completely assembled plant with all equipment installed on the concrete foundation provided by the purchaser. Model R "Oxigests" can be installed above ground (left) or below ground (right).

Write for complete engineering data manual on Smith & Loveless sewage treatment plants.
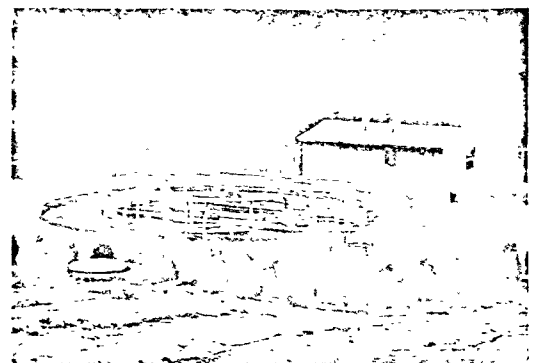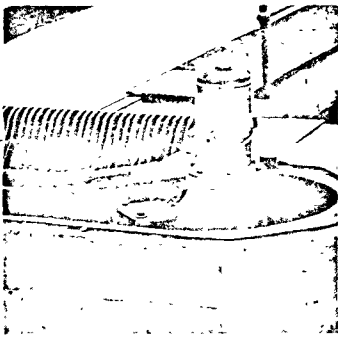
COMMINUTORS

Smith & Loveless offers a complete line of high-quality comminutors, engineered to provide continuous, automatic screening and cutting of large sewage solids into small particles without removing the solids from the sewage flow.

The comminutor is ideal for location at a sewage treatment plant, at sewage outfalls or for special applications at factories or commercial establishments—where ever large sewage solids must be screened, shredded or cut up. Smith & Loveless' complete line of comminutors are backed by more than 20 years' experience and proved by installations all over the world. For information on the quality comminutors offered by Smith & Loveless, write for the 100-page data manual on sewage treatment plants.
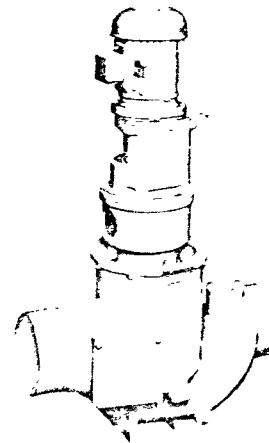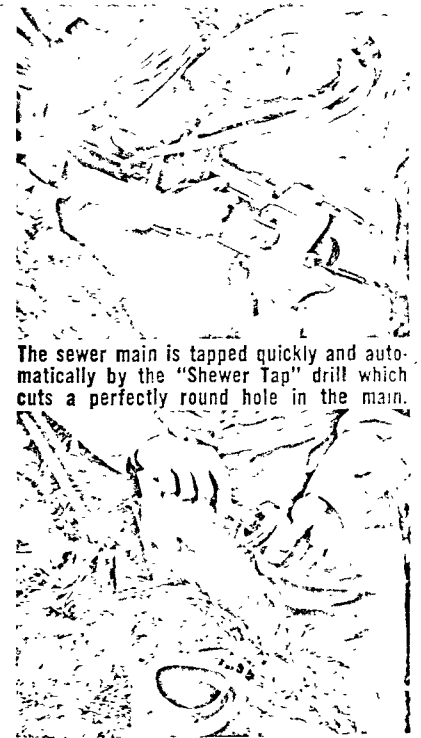
SEE BULLETIN 900

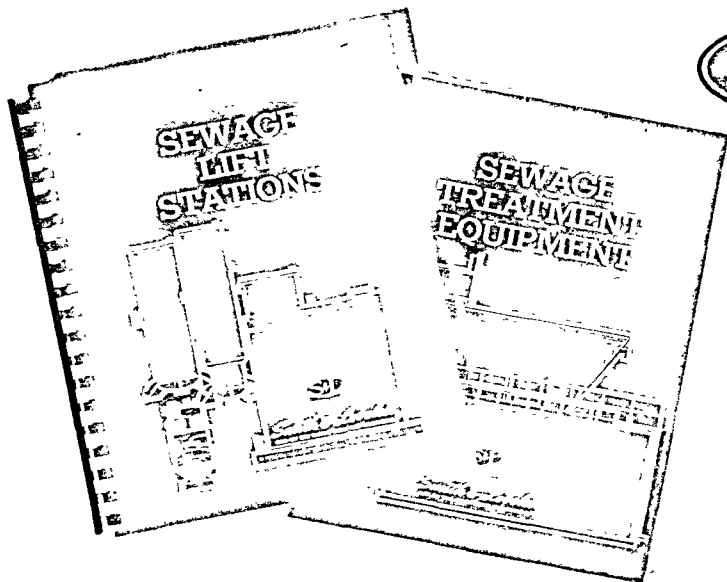# SHEWER TAP
METHOD OF TAPPING SEWER

The sewer main is tapped quickly and automatically by the "Shewer Tap" drill which cuts a perfectly round hole in the main.

Epoxy resin joint material is provided for each "Shewer Tap" connection.

The "Shewer Tap" method for tapping sewer mains to make house-to-sewer service connections is a fast, effective joining method with structural strength as strong as the pipe to which it is applied. It provides an absolutely water-tight, root-tight connection to eliminate the major cause of infiltration in sewers. This proven, effective sewer-tapping method is already being used by cities all across the nation. Write for complete information on the "Shewer Tap" method.
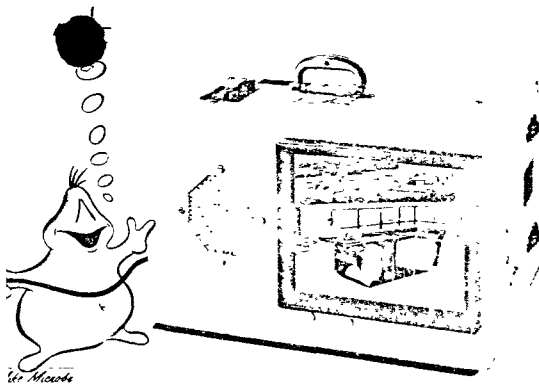
WRITE

for free engineering data manuals on Smith & Loveless sewage lift stations and sewage treatment plants. The manuals contain complete installation and operation data, selection and capacity charts, sample specifications, accessory equipment and dimension drawings.

Smith & Loveless' 16mm, color-sound industrial motion pictures on factory-built sewage lift stations and sewage treatment plants are available for viewing at meetings and conferences or in your own office via a unique, portable projector. Operational characteristics are explained by complete animated sequences. The movies show the design features, manufacture and installation of Smith & Loveless products.

Ask for the name of the Smith & Loveless representative in your area.

Smith & Loveless

Main Plant: Lenexa, Kansas

ONE OF THE
UNION
TANK CAR
COMPANIES

Factories:

Oakville, Ontario, Canada
Glasgow, Scotland
Brussels, Belgium

In U. S.: Smith & Loveless, Division-Union Tank Car Co., Lenexa, Kansas