

SESSION II

TERTIARY TREATMENT FOR WASTEWATER REUSE

**ADVANCED WASTEWATER TREATMENT
DESIGN SEMINAR
RIVERSIDE, CALIFORNIA
MARCH 24, 1972**

**U.S. ENVIRONMENTAL PROTECTION AGENCY
NATIONAL ENVIRONMENTAL RESEARCH CENTER
ADVANCED WASTE TREATMENT RESEARCH LABORATORY
CINCINNATI, OHIO**

Program for
Advanced Wastewater Treatment
Design Seminar

Riverside, California
March 24, 1972

Session Introduction

Mr. John Merrell, Director Categorical Programs Division EPA - Region IX San Francisco, California	30 min.	Introduction
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Session I - Control of Nitrogen in Wastewater Effluents

<u>Speaker</u>	<u>Time</u>	<u>Topic</u>
Mr. Edwin F. Barth Biological Treatment Research Prog. Environmental Protection Agency NERC-Cincinnati, Ohio	45 min.	Nitrogen control - general considerations including design of two sludge systems
Mr. Dennis Parker Brown & Caldwell, Inc. Consulting Engineers San Francisco, California	45 min.	Design of Central Contra Costa plant for nitrogen control
Coffee Break	15 min.	
Mr. John M. Smith Municipal Treatment Research Prog. Environmental Protection Agency NERC-Cincinnati, Ohio	60 min.	Upgrading existing facilities for nitrogen control including denitrification
Dr. Robert B. Dean, Chief Ultimate Disposal Program Environmental Protection Agency NERC-Cincinnati, Ohio	30 min.	Physical-chemical methods of nitrogen removal
	30 min.	Questions and discussion

Session II - Tertiary Treatment for Wastewater Reuse

<u>Speaker</u>	<u>Time</u>	<u>Topic</u>
Mr. A. N. Masse, Chief Municipal Treatment Research Prog. Environmental Protection Agency NERC-Cincinnati, Ohio	30 min.	Tertiary clarification
Dr. Robert B. Dean, Chief Ultimate Disposal Program Environmental Protection Agency NERC-Cincinnati, Ohio	45 min.	Lime recovery
Coffee Break		
Mr. George M. Wesner, Engineer Orange County Water District Orange County, California	45 min.	Clarification and filtration design
Staff Engineer - County Sanitation Districts of Los Angeles County	60 min.	Disinfection of municipal wastewater
	30 min.	Questions and discussion

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* PHOSPHORUS REMOVAL BY LIME TREATMENT OF SECONDARY EFFLUENT

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* FILTRATION

* DISINFECTION

COAGULATION OF WASTEWATER

CHAPTER 4

COAGULATION OF WASTEWATER

4.1 General

Wastewaters contain a wide variety of organic and inorganic suspended solids. These solids may have been present in the raw wastewater or may have been precipitated from solution during previous stages of treatment. All must be removed if a high quality effluent is to be produced.

A number of factors will influence the rates at which wastewater solids may be removed by sedimentation. Particularly important is particle size. Small particles in the colloidal range will not settle out in a practical detention time and must be agglomerated into larger particles which will settle at reasonable rates.

Small particles have very large surface areas per unit weight of solid with associated forces which tend to keep the particles separated or in a stable state. Effecting aggregation of the solids is a matter of overcoming these stabilizing forces by application of selected chemical coagulants. It is not the purpose of this manual to cover the subject of coagulation in detail. Only such information which is useful in process control is provided. More extensive coverage can be found in a number of publications for those who desire further information (1),(2), (3).

The principle stabilizing forces are electrostatic repulsion from electrical double layers surrounding particles suspended in water and physical separation of particles by films of water adsorbed on the particle surfaces. The electrical double layer is due to an imbalance of ions near the particle surface and imparts an electrical charge to particles which is generally negative for domestic wastewater particles (4). The particles with like charges then tend to repel one another. The magnitude of the charge on the double layer must be reduced to permit the particles to come together for agglomeration. Reduction of the charge is a principle function of the coagulant.

Principal natural destabilizing or aggregating forces are Brownian movement and Van der Waals forces. Brownian movement is a constant random motion of small particles due to collisions of the particles with thermally-agitated water molecules, while Van der Waals forces are atomic dipole interactions which exist between all atoms. Van der Waals forces, which are always attractive, predominate at short distances from the particle surfaces. If Brownian movement causes two or more particles to approach within a short distance from each other, and if the repulsive electrical forces have been sufficiently reduced, then the particles will be held together by the Van der Waal forces. Other particles can be added to the particle pair until the resulting aggregate or floc reaches such a size that it rapidly settles out of solution.

These natural destabilizing forces are generally not sufficiently rapid to permit effective solids removal from wastewater, but must be augmented by addition of chemical coagulants to destabilize and tie the particles together and by application of hydraulically-or mechanically-applied mixing to promote rapid collisions of the destabilized particles. A number of mechanisms have been proposed to explain the action of coagulants in promoting particle

aggregation. These include charge reduction, physical or chemical bridging of coagulant molecular chains between particles and physical enmeshment of particles in a mass of precipitated coagulant. All of these mechanisms probably play a greater or lesser role in effective coagulation.

Coagulants used in wastewater treatment include those used in potable water treatment with some additions. These are: alum, sodium aluminate, ferric chloride, ferric sulfate, ferrous chloride, ferrous sulfate, lime and organic polyelectrolytes. Other materials such as soda ash or clays may be used as sources of alkalinity or weighting agents, respectively, to aid coagulation. When added to water, salts of aluminum or iron react with the water or alkalinity present in the water to form insoluble hydrolysis products. It is these hydrolysis products which are the effective coagulating agents. These materials which are positively charged in the neutral pH range adsorb on the negatively charged wastewater particles reducing repulsive forces between the particles. The coagulant may also react with other constituents of the wastewater, particularly anions such as phosphate and sulfate, forming hydrolysis products containing various mixtures of ions. These various hydrolysis products differ in their effectiveness as coagulants. The chemistry of the reactions is extremely complex.

For each combination of coagulant and wastewater there is an optimum dosage of coagulant and an optimum pH range for coagulation. These are the two parameters which are generally controlled in operation of advanced waste treatment plants for solids removal through coagulation.

4.2 Coagulation Control

Because coagulation represents a group of complex reactions, laboratory experimentation is essential to establish and maintain the optimum coagulant dosage and the effect of important variables on the quality of coagulation of the wastewater under investigation. With hydrolyzing coagulants three procedures may be followed for this purpose: the jar test, measurement of zeta potential, and measurement of phosphate content of the wastewater. Lime is a special case. Proper control of lime addition may usually be maintained by measuring the pH or automatically titrating alkalinity after lime addition.

4.2.1 Jar Test

The single, most widely used test to determine dosage and other parameters is the jar test. The equipment for this test and the directions for its proper performance have been published(5),(6),(7),(8). The jar test attempts to simulate the full scale coagulation-flocculation process and has remained the most common control test in the laboratory since its introduction in 1918. Since the intent is to simulate an individual plant's conditions, it is not surprising that there has been no standardization of the test. The jar test as variously performed does, however, have some elements of conformity. In its essentials, the jar test simply consists of a series of sample containers, usually six, the contents of which can be stirred by individual mechanically-operated stirrers. Water to be treated is placed in the containers and treatment chemicals are added while the contents are being stirred. The range of conditions, for example, coagulant dosages and pH, are selected to bracket the anticipated optima. After a short, 1-5 minute, period of rapid stirring to ensure complete dispersion of coagulant, the stirring rate is decreased and flocculation is allowed to continue for a variable period, 10 to 30 minutes or more, depending on the simulation. The stirring is then stopped and the floc are allowed to settle for a selected time. The supernatant is then

analyzed for a variety of parameters. With wastewater the usual analyses are for turbidity or suspended solids, pH, residual phosphorus and residual coagulant.

If desired, a number of supernatant samples may be taken at intervals during the settling period to permit construction of a set of settling curves which provide more information on the settling characteristics of floc than a single sample taken after a fixed settling period. A dynamic settling test may also be used in which the paddles are operated at 2 to 5 rpm during the settling period. This type of operation more closely represents settling conditions in a large horizontal basin with continuous flow.

Several six-position stirrers are available commercially for running jar tests; one from Phipps and Bird, (Phipps and Bird, Inc., Richmond, Va.), another from Coffman Industries, (Coffman Industries, Inc., Kansas City, Ka.), are shown in Figure 4-1. Standard laboratory mixers have also been used; however, it is difficult to obtain reproducible mixing conditions using different pieces of equipment. Various types of containers, usually beakers or jars, are used to hold the samples. Improved mixing may be obtained by adding stationary plates in the containers as described by Camp and Conklin (8). The Coffman stirrer has an attachment which makes it possible to add coagulant to all containers simultaneously, however, good results can be obtained by rapidly adding coagulant from a large graduated pipette to each jar in sequence.

A simple apparatus, shown in Figure 4-2, can be constructed from tubing, rubber stoppers and small aquarium valves to permit rapid sampling of supernatant. The unit is placed next to the sample jars at the beginning of the settling period with the curved stainless steel tubes dipping into the jars. At desired intervals the vent valve is covered with a finger, permitting vacuum to draw samples into the small sample bottles. The needle valves are adjusted so that supernatant is drawn into all the bottles at the same rate. When sufficient sample is obtained, the vent is uncovered and the bottles are replaced with empties. The maximum sampling rate is about once per minute.

Figure 4-3 shows typical types of settling curves which may be obtained. Curve A indicates a coagulation which produced a uniformly fine floc, so small that at the end of 1 to 2 minutes settling, the supernatant had a turbidity equal to that of the starting water due, in part, to the fine floc which resisted settling. Settling was slow and the final turbidity was excessive. This coagulation would not be satisfactory in an advanced wastewater treatment plant. Curve B represents the most common type of settling rate obtained. During the first 5 minutes, the settling rate was practically a straight line on a semilog plot. Settling was rapid and clarification was satisfactory. The coagulation represented by curve C shows that a mixture of large rapid settling floc and small, slow-settling particles was obtained. Settling was rapid for the first two minutes, but with little further clarification after that. High residual turbidity may also have resulted from incomplete coagulation. Curve D represents the ultimate in coagulation. Practically all of the floc particles were so large and dense that 97% settled within three minutes. Sedimentation was essentially complete within that time since only 0.5% additional floc settled in the next 27 minutes. Final clarity of the supernatant was entirely satisfactory. This coagulation was obtained with a coagulant aid.

Measurement of turbidity provides the most rapid indication of the degree of solids removal obtained. The recommended procedure for turbidity measurement by light scattering is given in the 13th edition of *Standard Methods for Examination of Water and Wastewater*; however, other methods varying from simple visual evaluation to measurement of light

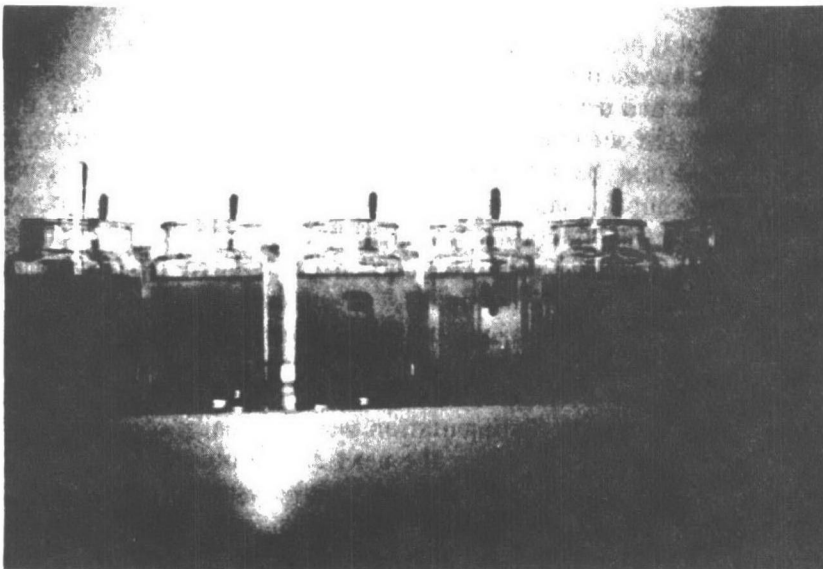
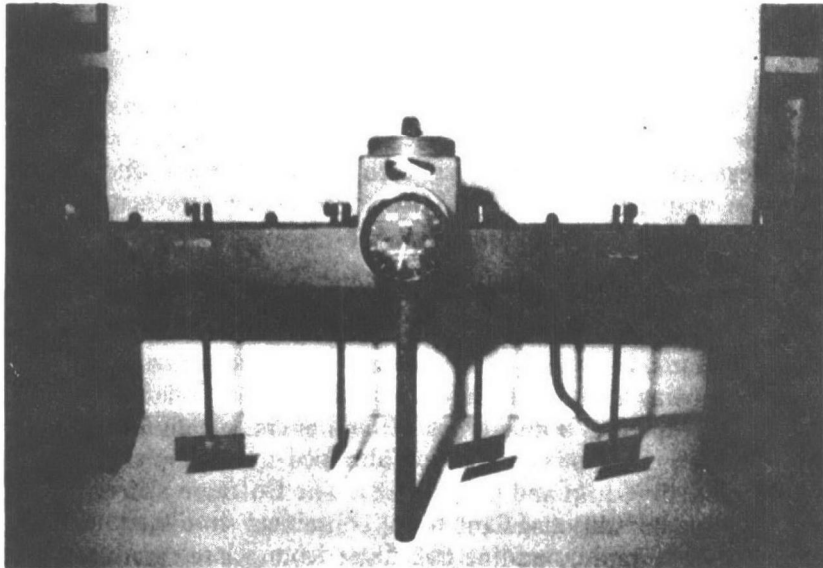


Figure 4-1. JAR TEST UNITS WITH MECHANICAL (TOP)
AND MAGNETIC (BOTTOM) STIRRERS

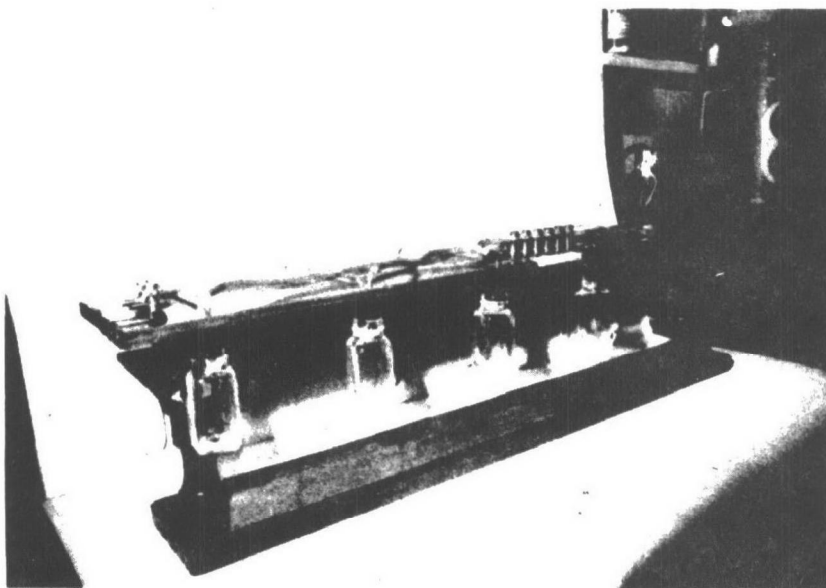
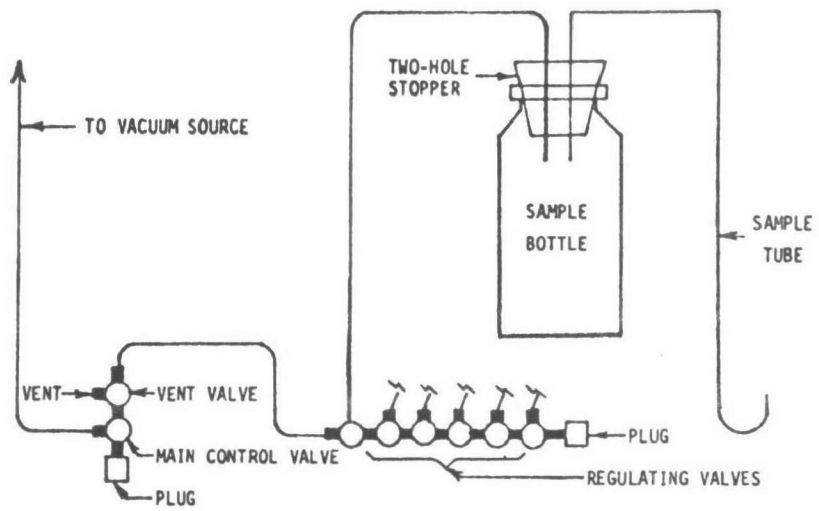


Figure 4-2. SIX-POSITION SAMPLER

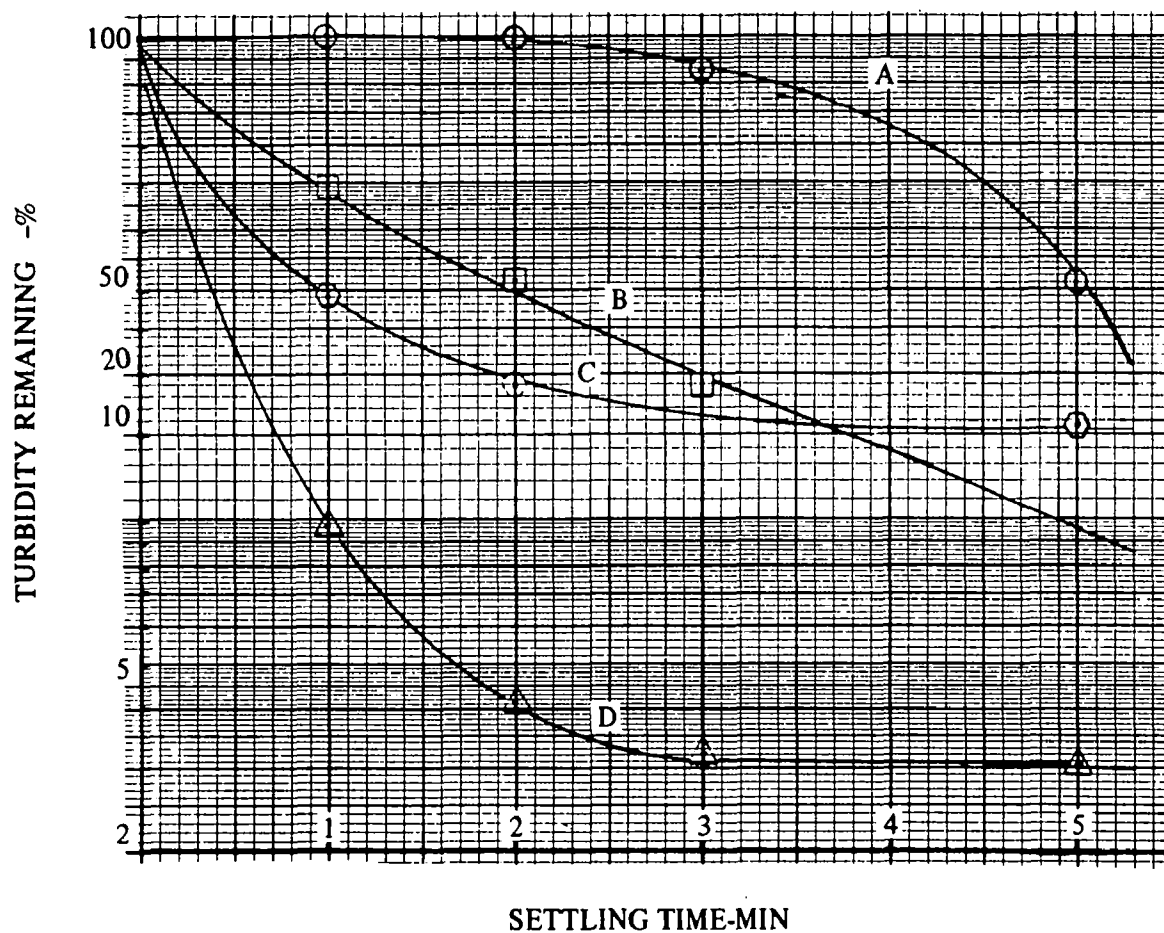


Figure 4-3. SETTLING CURVES FREQUENTLY OBTAINED

transmitted on a laboratory spectrophotometer can be used for purposes of comparison. Measurement of residual suspended solids is the only procedure which gives the actual weight concentration of solids remaining, but the procedure is too slow for purposes of process control. Residuals of phosphate and coagulant in supernatant are usually of interest and may be measured either manually or with automated equipment.

A typical jar test might be run as follows: Wastewater samples are placed in containers and rapid mix is started at 100 rpm. Selected dosages of coagulant are rapidly added to the containers covering the expected range of the optimum dosage and a timer is started. If a polymer is to be used as a coagulant aid, it is added to each jar after two minutes and rapid mixing is continued for one additional minute. The paddles are then slowed to 30 rpm and mixing continues for 10 minutes. The paddles are then stopped and the sampling apparatus previously-described is placed in position. At settling times of 1, 3, 5, 10 and 20 minutes samples of supernatant are drawn for turbidity measurement. After the final turbidity sample is drawn, a larger volume of supernatant may be decanted for more complete analysis. Results are plotted as in Figure 4-4 for judgment as to the desired coagulant dosage. The jar test may be repeated using a smaller dosage range around the observed optimum to more closely locate the best dosage.

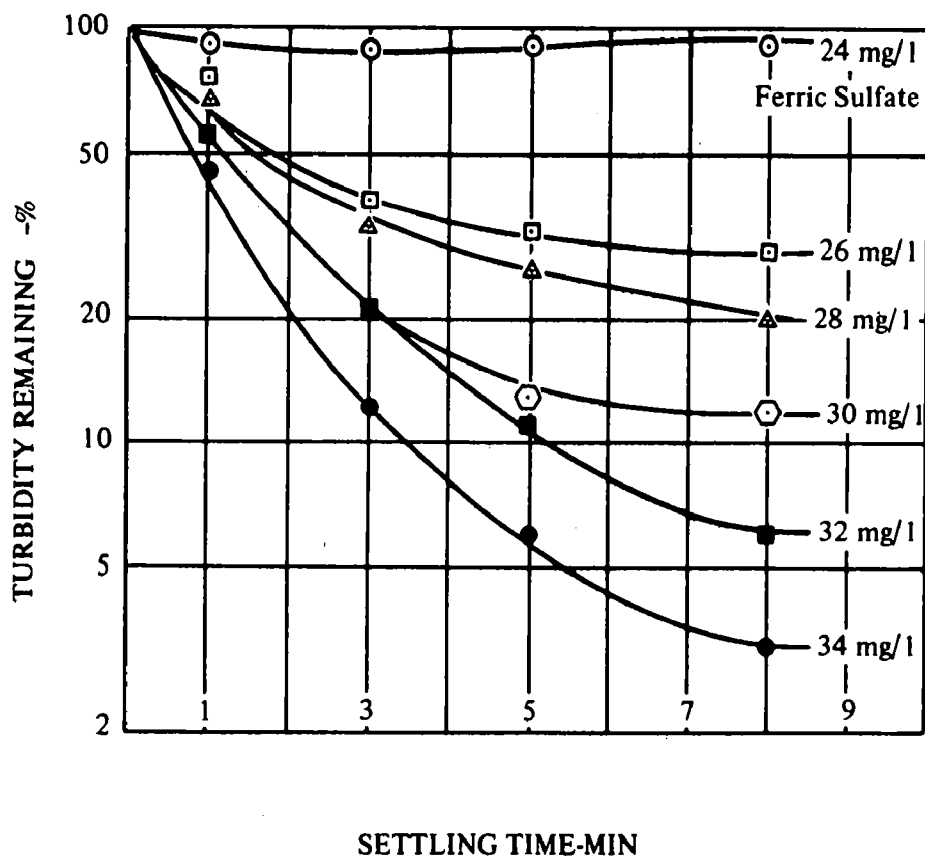


Figure 4-4. JAR TEST RESULTS

If additional alkalinity is required to hold the coagulation in the optimum pH range, this should be added to the samples ahead of the coagulant. Once an approximate optimum dosage has been determined, it may be desirable to repeat the jar test with the optimum coagulant dosage but using varying quantities of added alkalinity to give different pH values. Experience in coagulating a given wastewater provides the best guide as to the best methods for controlling the process.

4.2.2 Zeta Potential

Measurement of particle charge is another procedure which may be useful for control of the coagulation process(9),(10),(11). The total particle charge is distributed over two concentric layers of water surrounding the particle; an inner layer of water and ions which is tightly bound to the particle and moves with it through the solution and an outer layer which is a part of the bulk water phase and moves independently of the particle. The surface charge is not measurable experimentally; however, the zeta potential which is the residual charge at the interface between the layer of bound water and the mobile water phase can be determined with a commercially-available instrument.*

In the zeta potential measurement procedure, a sample of treated water containing floc is placed in a special plastic cell under a microscope as shown in Figure 4-5. Under the influence of a voltage applied to electrodes at the ends of the cell, the charged particles will migrate to the electrode having a polarity opposite that of the particle. The velocity of migration will be proportional to the particle charge and to the applied voltage. The particle velocity can be calculated by observing the time it takes a particle to travel a given distance across an ocular micrometer. The zeta potential can then be obtained from a chart which combines the particle velocity with instrumental parameters. Detailed operating instructions are supplied with the instrument.

To control the coagulation by zeta potential, samples of water while being mixed are dosed with different concentrations of coagulant. Zeta potentials are then measured and recorded for floc in each sample. The dosage which produces the desired zeta potential value is applied to the treatment plant. Zeta potentials of floc produced in the plant may also be measured as a means of control. The precise zeta potential which signals optimum coagulation must be determined for a given wastewater by actual correlation with jar tests or with plant performance as in Figure 4-6. The control point is generally in the range of 0 to -10 millivolts. If good correlations can be obtained between some zeta potential values and optimum plant performance, then it is possible to make rapid measurements of particle charge to compensate for major variations in wastewater composition due to storm flows or other causes. Short term variations such as those due to sudden industrial waste dumps are still beyond control with any present techniques because of the time lag between recognition of a problem with coagulation and adoption of a satisfactory change of coagulation conditions.

4.2.3 Phosphate Monitoring

A third means of coagulant dosage control where the coagulant is being used to precipitate phosphate as well as to remove solids is to automatically analyze the incoming wastewater for soluble orthophosphate. The coagulant is then paced to maintain a selected ratio of

*A product of Zeta-Meter, Inc., N.Y., N.Y.

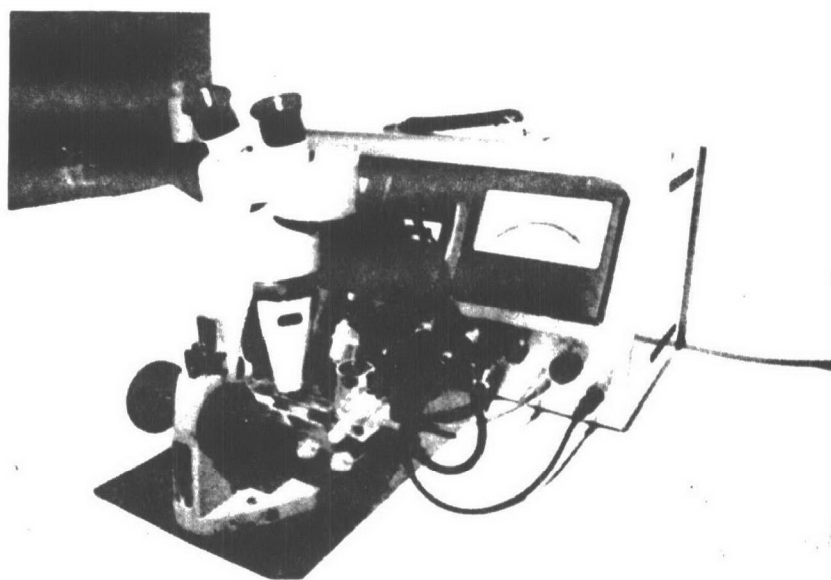


Figure 4-5. ZETA POTENTIAL APPARATUS

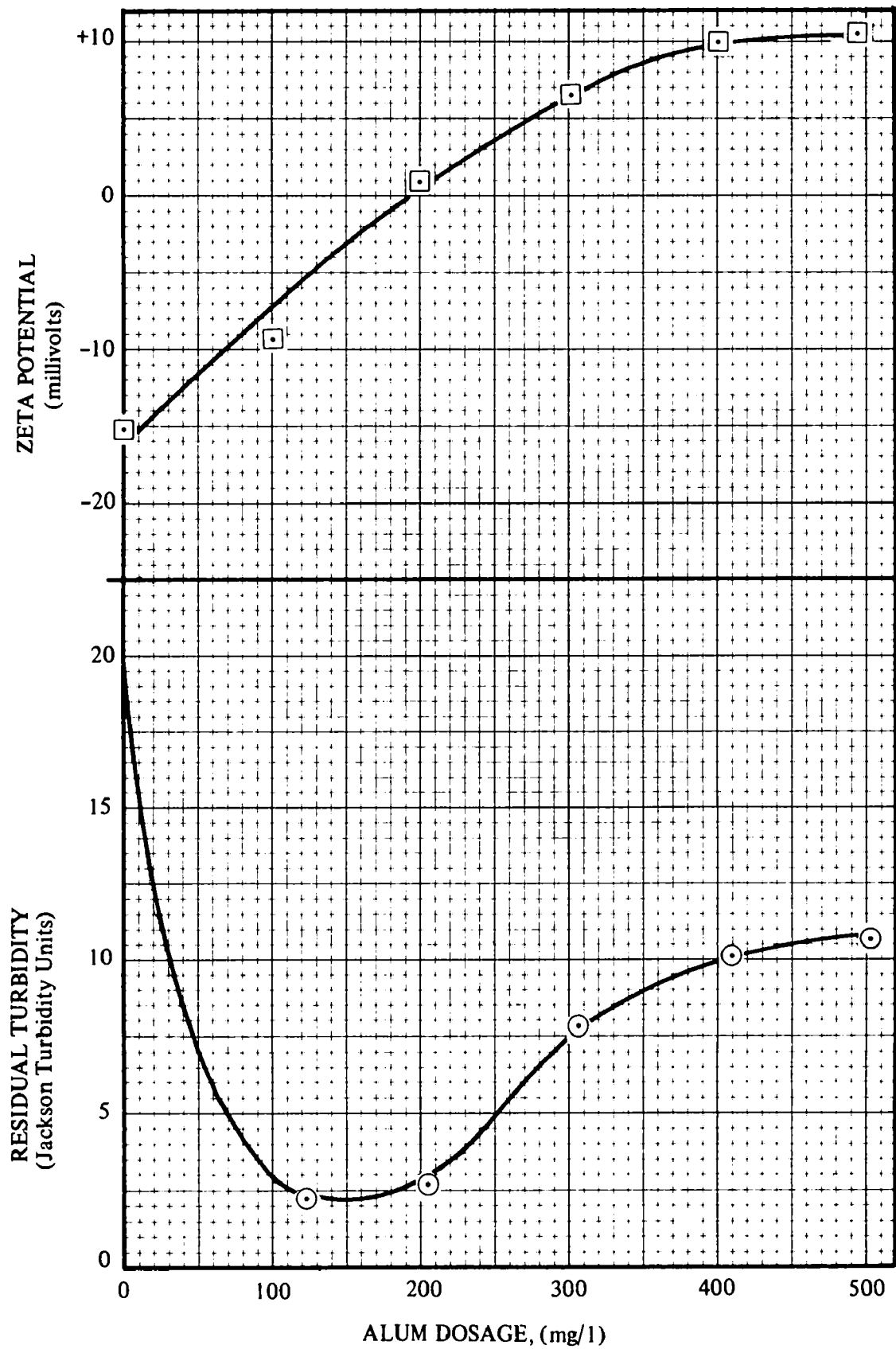


Figure 4-6. COAGULATION OF RAW SEWAGE WITH ALUM

coagulant to phosphate either automatically or by frequent manual adjustment. The Technicon Auto-Analyzer* which is commercially available has been adapted for this purpose. Dow Chemical Co. has developed an automatic system which will add ferric chloride in proportion to soluble orthophosphate with corrections for varying flow and concentration of ferric chloride.

Coagulant dosages which produce good phosphate removal will also generally result in good solids removal if polymers are used to aid in flocculating the fine precipitated matter. Under normal conditions, the polymer feed rate should be varied with flow to maintain a constant dosage; however, the polymer feed rate does not have to be changed with changes in dosage of the inorganic coagulant.

**A product of Technicon Corporation, Tarrytown, N.Y.*

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11. Riddick, T.M., Control of Colloid Stability Through Zeta Potential, Vol. 1, Zeta-Meter, Inc., 1720 First Avenue, New York, N.Y. 10028.

PHOSPHORUS REMOVAL BY LIME TREATMENT OF SECONDARY EFFLUENT

Chapter 8

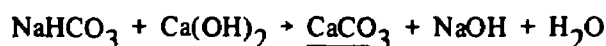
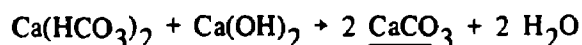
PHOSPHORUS REMOVAL BY LIME TREATMENT OF SECONDARY EFFLUENT

8.1 Description of Process

8.1.1 Theory

Lime treatment of wastewater is essentially the same process as the familiar lime softening of drinking water supplies. The objectives, however, are quite different. While softening may occur, the primary objective is to remove phosphorus by precipitation as hydroxyapatite. This reaction was described in Chapter 3.

During phosphorus precipitation other important reactions occur. The reaction of lime with alkalinity, that results in calcium removal when carrying out softening, not only takes place when treating wastewater, but may have a very important effect on the general efficiency of the process. This reaction can be considered to take place in the two following ways:



The first equation is that for softening. Some wastewaters do not contain enough calcium, however, for that equation to be satisfied. Calcium carbonate precipitation may still occur, but by the second reaction. The reactions to form CaCO_3 are important for two reasons: the lime consumption determines to a considerable extent the lime dose required for operating the process, and the resulting CaCO_3 acts as a weighting agent to aid in settling of sludge.

Another reaction that may be important is precipitation of $\text{Mg}(\text{OH})_2$ as follows:



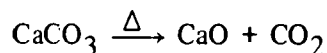
This reaction does not approach completion until the pH is raised to 11. Magnesium hydroxide is a gelatinous precipitate which aids colloid removal, but which hinders sludge thickening and dewatering. Where the pH must be raised to 11 or above to meet a phosphorus removal requirement or some other treatment objective, $\text{Mg}(\text{OH})_2$ formation must be considered.

In the two-stage lime treatment process which will be described below, it is necessary to include recarbonation after the first stage to reduce the pH and precipitate the excess lime as CaCO_3 in the second stage. The following reaction occurs:



Carbon dioxide may also be used to lower pH after lime treatment. The important reaction in this case would be the conversion of CO_3^{2-} to HCO_3^- .

A final reaction that is of major concern, where lime is to be recovered by sludge recalcination, is as follows:



The CaO produced would then be slaked to form Ca(OH)_2 before use.

8.1.2 Treatment Systems

Two lime treatment systems may be used with wastewater, single-stage and two-stage. Figure 8-1 shows a single-stage system. In single-stage treatment lime is mixed with feed water to raise the pH to a desired value. Although the pH will depend upon the required phosphorus removal, it is likely to be substantially less than 11, and may be less than 10. Precipitation of phosphate and other materials, as indicated by reactions discussed above, takes place. Time is allowed in the appropriate equipment for precipitate particles to flocculate to sufficient size for good settling. The clarified water from the settler may be discharged directly or may be filtered to improve solids removal. Adjustment of pH with CO_2 may be necessary before discharge and will almost certainly be required before filtration to prevent post-precipitation of CaCO_3 from the unstable water. The settled lime sludge may be disposed of as landfill or may be recalcined for recovery of lime. In the latter case, the sludge is thickened, dewatered by centrifuge or vacuum filter, and calcined. The calcined product is then slaked and reused. To avoid buildup of inerts in the lime, some of the sludge or recalcined lime must be wasted or the inerts must be separated from the sludge before calcination.

Two-stage treatment is somewhat more complicated than single-stage treatment. In typical two-stage treatment, shown in Figure 8-2, enough lime is added to the water in the first stage to raise the pH above 11. Precipitation of hydroxyapatite, CaCO_3 , and Mg(OH)_2 occurs. Consideration of the solubility product for CaCO_3 and the equilibrium between CO_3^{2-} and HCO_3^- shows that the minimum solubility for Ca^{2+} occurs at a pH of about 10. At a pH of 11 or above there is a considerable concentration of Ca^{2+} present in the water. In two-stage treatment CO_2 is added after the first-stage settler to bring the pH down to about 10 where CaCO_3 precipitation results. The CaCO_3 is settled out and the clarified water is either discharged or sent to filtration. As in the case of single-stage treatment, pH reduction may be necessary before discharge and probably would be required before filtration.

8.2 Typical Performance Data

A number of full-scale and pilot-scale tertiary lime treatment plants are in operation and more full-scale plants are beginning operation. Only one plant, that at South Lake Tahoe, California, has operated for a significant period with recalcination of sludge for lime recovery. Most aspects of this 7.5 mgd plant have been discussed in detail by Culp and Culp (1). Although the data on handling of wastewater sludges for recalcination and the recalcination process itself are limited mainly to this plant,

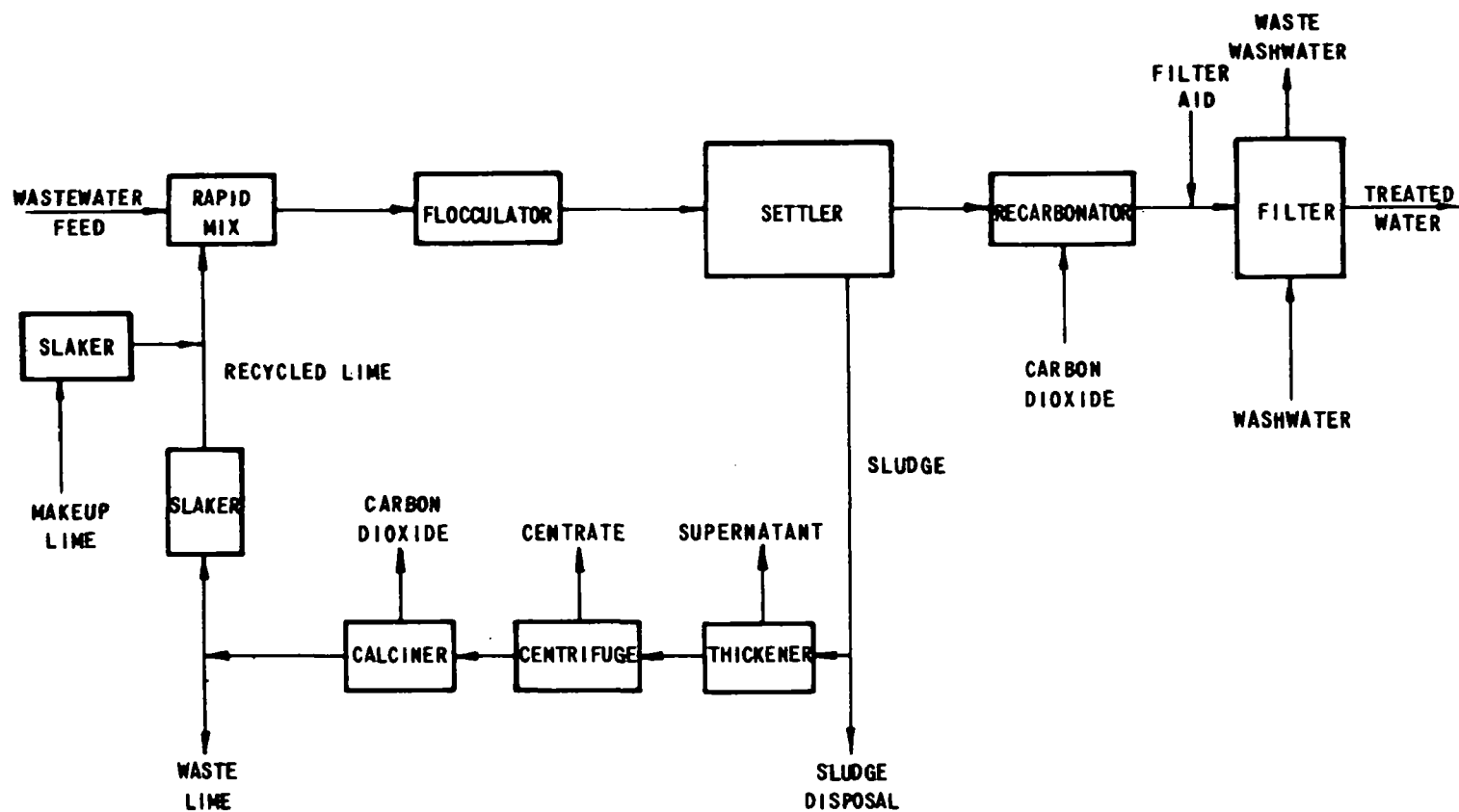


FIGURE 8-1 SINGLE STAGE LIME TREATMENT SYSTEM

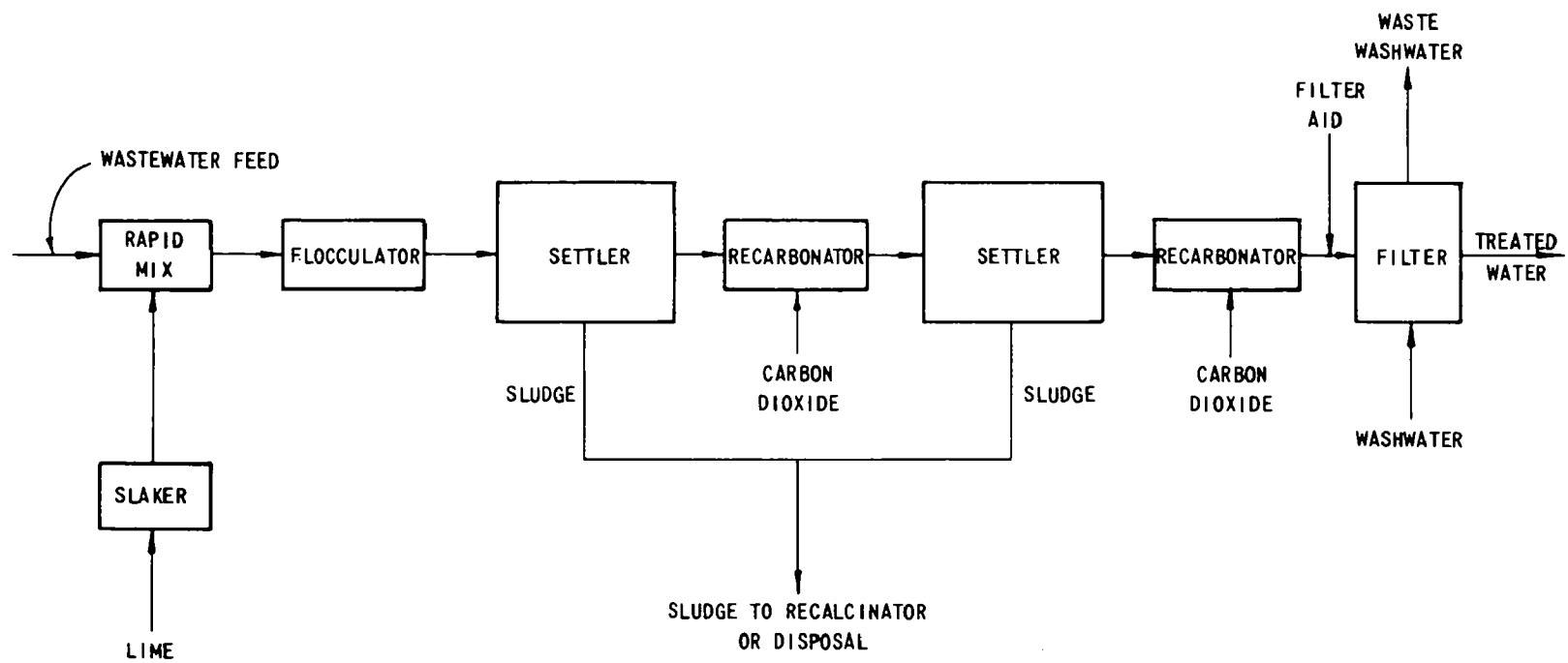


FIGURE 8-2 TWO STAGE LIME TREATMENT SYSTEM

performance data on other parts of a lime treatment system are available from more than one location. Extensive data have been reported from pilot plants located in Washington, D. C., (2) and Lebanon, Ohio (3).

The lime treatment system at Tahoe is a two-stage clarification system which is followed by pressure filters of the multimedia type. These contain 3 ft. of a mixture of coal, sand, and garnet. The water passes through two filters in series. The first stage clarifier is operated at a pH of 11. To reach that pH requires about 300 mg/l CaO. A small amount of polymer is used to improve flocculation. The pH is reduced to 9.6 by recarbonation before the second-stage settler. Before filtration, the pH is further reduced to 7.5. A small amount of alum (1 to 20 mg/l) or a combination of alum and polymer is used as filter aid. Filter run length has varied between 4 and 60 hours.

Phosphorus removal at this plant always has been good and has improved as operating experience has increased. By returning plant waste streams containing precipitated phosphorus to the first stage flocculator, it is now possible to obtain routinely an effluent with less than 0.1 mg/l P. Before the filters the phosphorus concentration is about 0.4 mg/l.

In addition to phosphorus removal, there is significant removal of organic materials. During a period of intensive study of the first stage of clarification (unpublished data) 77% removal of BOD and 61% removal of COD were obtained. Although there was not a large removal of suspended solids, there was a significant change in the character of the solids from organic to largely inorganic. Further removal of organic materials and suspended solids occurs over the remainder of the system. Culp and Culp report typical filter effluents with a BOD of 3 mg/l, COD of 25 mg/l, and turbidity of 0.3 JTU.

Sludges from the first stage settler and the settler following recarbonation are sent to a gravity thickener. Solids concentration increases during thickening from about 1% to from 8 to 20%. Thickened sludge is then centrifuged to form a cake with from 30% to more than 40% solids. The cake is next calcined in a multiple hearth furnace and the recovered lime slaked for reuse. Initially it was planned to operate the centrifuge for high solids recovery in the cake. This necessitates discarding a significant amount of the recovered lime to prevent buildup of precipitated phosphate and other inerts. Since operation began, it has been found that much of the phosphate and some other inert materials can be separated from the CaCO_3 in the sludge by operating with a rather high solids concentration in the centrate. Approximately 90% of the phosphorus can be removed in the centrate, for example, when 25% of the solids entering the centrifuge are allowed to remain in that stream. This classification procedure results in a loss of about 15% of the recoverable lime. A greater degree of utilization of recalcined lime compensates for the loss, however, and the load on the calcining furnace is reduced. Over three years of operation, mostly without classification in the centrifuge, the average concentration of CaO in the recalcined product was 66%. Recalcined lime made up 72% of the total used at the plant.

Considerable operating data have been obtained by O'Farrell and Bishop (2) on a two-stage system operated at the EPAWQO-DC Pilot Plant in Washington, D. C. This 50,000 gpd plant treated secondary effluent from a pilot activated sludge system at the municipal treatment plant. The first stage pH was maintained at about 11.7, using a lime dose of about 400 mg/l as CaO, and the pH after recarbonation was maintained at about 10.3. Ferric iron was added at a rate of 5 mg/l in the second stage to improve flocculation. Water from the second-stage settler was filtered without further pH adjustment through gravity-flow dual-media filters consisting of anthracite coal over sand. The average filter run length was more than 50 hours.

Phosphorus removal for the system was similar to that obtained at Tahoe; 0.09 mg/l as P was the average concentration remaining. Phosphorus concentration before filtration averaged 0.13 mg/l.

In addition there was significant organic and suspended solids removal. The average BOD was reduced from 15 mg/l to 2.1 mg/l before and 1.5 mg/l after filtration. Suspended solids were reduced from 33 mg/l to 17 mg/l before filtration and 3.8 mg/l after filtration.

Sludge from the second stage settler was returned to the first stage and all sludge removed from the system was from the settler of the first stage. This sludge contained about 5% solids and constituted about 1.5% of the feed volume. Although there was sludge thickening and calcining equipment available, only preliminary study was made of lime reuse. Limited results showed improved thickening of the sludge when recalcined lime was recycled to the system.

A 75 gpm single-stage pilot system was operated at the Lebanon, Ohio Sewage Treatment Plant. Feed water was activated sludge effluent. This system, consisting of a flocculator-clarifier followed by dual-media filters of anthracite coal and sand, is described by Berg, et al. (3). Sludge was gravity thickened and discharged to sand drying beds. The system has been operated over a pH range from about 9 to 11 with most data taken at a pH of 9.5. Before filtration the pH was reduced to about 8.8 with sulfuric acid to prevent precipitation on the filter media. Filter run length averaged about 90 hours.

Effluent phosphorus concentrations for the system are shown in Figure 8-3. The effect of pH is clearly indicated. Since the average phosphorus concentration of the secondary effluent was about 10 mg/l as P, approximately 95% removal was obtained at a pH as low as 9.5. Before filtration the average phosphorus concentration was 0.75 mg/l at a pH of 9.5. Removal improved only slightly at higher pH because of the presence of precipitated phosphorus in the effluent.

Average suspended solids concentration of the secondary effluent was 43.5 mg/l. This was reduced to 16.5 mg/l in the clarifier. Much of the suspended matter in the clarifier effluent consisted of inorganic precipitates. After filtration the water was of high clarity. During a six month period when the biological treatment plant operated well, turbidity averaged 0.2 JTU. Even when the suspended solids load from the activated

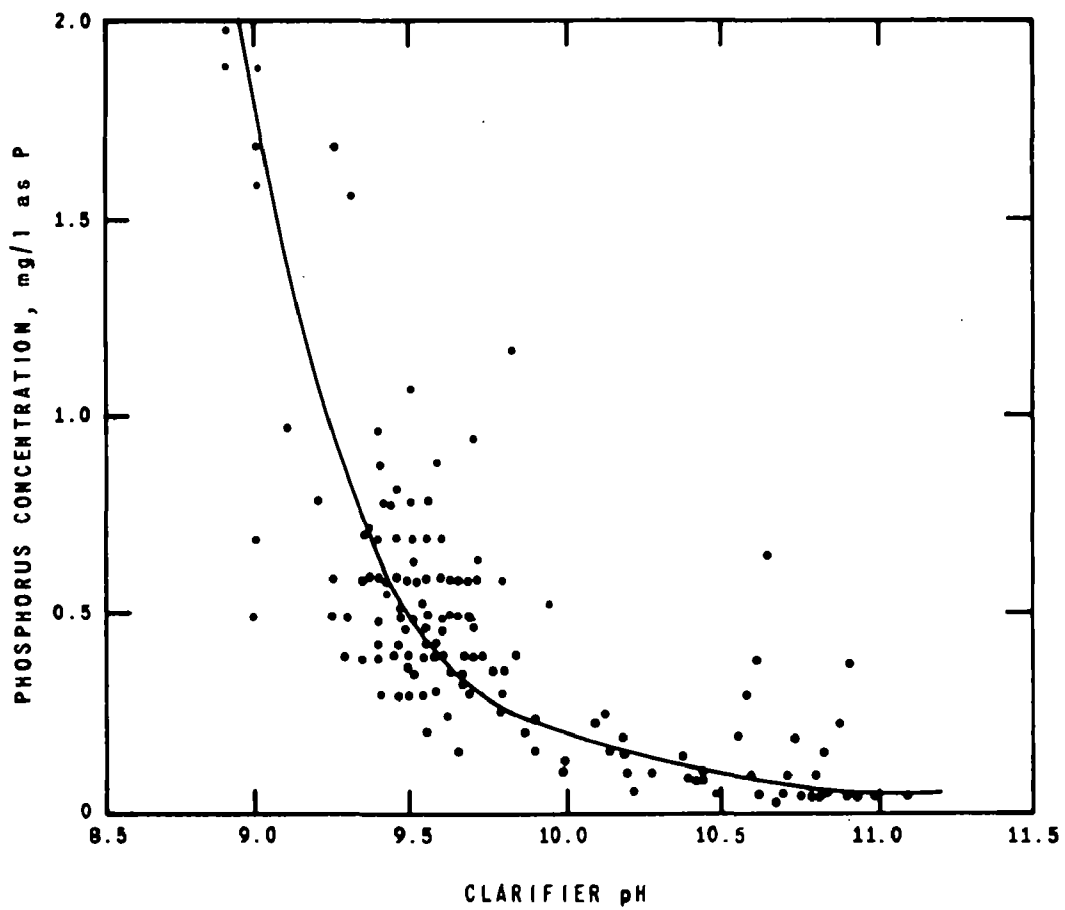


FIGURE 8-3 EFFECT OF pH
ON PHOSPHORUS CONCENTRATION
OF EFFLUENT FROM FILTERS
FOLLOWING LIME CLARIFIER

sludge plant was high, the lime treatment system produced water with a daily average turbidity not exceeding 2.5 JTU.

During a two month period in which extensive measurement of organic removal was made, total organic carbon was reduced by 63% over the clarifier and 68% over the clarifier and filters. Organic carbon measurements taken at other times indicated removals of from 55 to 74%. Occasional BOD and COD samples indicated removals of 86% and 62%.

Sludge was usually removed from the settler at 1.5 gpm giving a sludge concentration of 2.8% solids. This was thickened to about 10% before being pumped to drying beds. The sludge dewatered quickly to about 50% solids. Equipment was not available for recalcining the sludge.

The lime requirement is an important consideration in lime treatment. This can vary over a wide range depending on operating pH and water composition. From reactions given earlier it was seen that alkalinity has an important effect on the lime dose. Buzzell and Sawyer (4) have shown, for example, that for several wastewaters the lime dose required to reach a pH of 11 correlated approximately with alkalinity. Examination of the earlier equations would show also that calcium hardness can affect lime dose. One part by weight of CaO can react with from 0.89 to 1.79 parts of bicarbonate alkalinity expressed as CaCO_3 , the lower value applying to very soft waters and the higher value to very hard waters. In addition to the reaction of lime with hardness, other competing reactions occur in lime treatment of wastewater. Also, there may be incomplete reaction of the lime. All of these complications make calculation of lime dose difficult. The result is that, at present, determination of lime dose is largely empirical. Approximate values have already been given for the plants at Tahoe and Washington, D. C. Some approximate values for the Lebanon work are shown in Table 8-1. It appears from the data already available that the lime dose will usually be in the range of 300 to 400 mg/l as CaO for two-stage treatment, and from 150 to 200 mg/l where single-stage treatment is satisfactory.

Table 8-1

LIME REQUIREMENTS

<u>Feed Water</u> <u>Alkalinity</u> (mg/l as CaCO_3)	<u>Clarifier pH</u>	<u>Approximate Lime</u> <u>Dose</u> (mg/l of CaO)
300	9.5	185
300	10.5	270
400	9.5	230
400	10.5	380

8.3 Criteria for Selection of Process

Lime treatment of secondary effluent represents a significantly higher capital cost and somewhat higher operating cost for phosphorus removal than mineral addition to a conventional treatment plant. Lime treatment has, however, several advantages over mineral addition. It can be considered more dependable since it adds additional flocculation and sedimentation steps to the system. Upsets in the conventional plant, which would reduce the efficiency of phosphorus removal by mineral addition, would have less effect on tertiary lime treatment. Because tertiary lime treatment is separate from the conventional treatment plant, it adds flexibility to operation of the system. For very high degrees of phosphorus removal, lime treatment would be the method of choice, not only because of the inherent greater dependability mentioned above, but because of the ability to produce an effluent slightly lower in phosphorus content. Lime treatment decreases the total dissolved solids content of the water by removal of hardness and alkalinity while mineral addition adds to the total dissolved solids. In the case of alum addition, for example, 5.3 parts by weight of sulfate are added for each part of aluminum. Lime treatment has the capability of removing turbidity to very low levels. Where there are plans for recreational reuse or certain industrial reuses of the treated water, low turbidity along with low phosphorus content of the lime treatment effluent is very desirable. Lime treatment offers the opportunity for recovery of the treatment chemical. At the present time there are no acceptable methods for recovery of aluminum or iron salts.

The choice of single-stage or two-stage lime treatment depends partly upon the degree of phosphorus removal required, but more importantly, on the alkalinity of the water. Unless a high treatment pH is used, waters with low alkalinity, in the range of 150 mg/l as CaCO_3 or less, form a poorly settleable floc because of the low fraction of dense CaCO_3 . A pH above 10 is needed even to obtain measurable CaCO_3 production. A pH of 11 or above is then needed to precipitate $\text{Mg}(\text{OH})_2$ which aids in settling of fine particles. Since neither discharge nor reuse of the high-pH water is likely to be acceptable, addition of a second treatment stage after recarbonation to a pH of about 10 becomes generally necessary. It would be possible to recarbonate to a much lower pH and avoid the second treatment stage, but this would result in a high calcium effluent and would eliminate the chance to produce high CaCO_3 sludge in situations where lime recovery was contemplated. Although most of the phosphorus removal occurs in the high pH first stage, in accordance with the pH dependence of phosphorus solubility as shown earlier by Figure 8-3, some removal does also occur in the second stage. Another important reason for including the second stage is to assure better control of clarification. With low alkalinity waters there is sometimes difficulty in settling the sludge in the first stage even at high pH. The second stage settler with its high CaCO_3 sludge, prevents solids carryover when the first stage settler does not operate properly.

With high alkalinity waters, a well settling floc is formed at pH values as low as 9.5. There is no need for two-stage treatment unless the required degree of phosphorus removal necessitates a high pH. In addition to obtaining a small degree of phosphorus

removal, the second stage would be used in these cases to lower calcium content in the effluent and to obtain high CaCO_3 sludge for lime recovery. There is not yet available operating experience at enough plants to state positively the alkalinity at which single-stage treatment will perform satisfactorily, from the standpoint of floc settleability. At present, experience indicates that at an alkalinity of 150 mg/l as CaCO_3 , single-stage treatment probably cannot be used. Between 150 mg/l and 200 mg/l settleability will probably depend on the amount of organic floc present. Above 200 mg/l, settleability is likely to be satisfactory.

8.4 Description of and Criteria for Choice of Equipment

8.4.1 Single-Stage System and First Stage of a Two-Stage System

Diagrams for single and two-stage lime treatment systems are shown in Figures 8-1 and 8-2. The first part of each system is made up of clarification equipment essentially the same as that found in water treatment plants. It includes a chemical feeding system (see Chapter 10), a rapid mix tank, a flocculator, a settler, and a means for sludge removal. Mixing, flocculation, and settling may be carried out in separate vessels or tanks, or they may be combined in one integrated unit. The system consisting of separate tanks has been used for many years in water treatment and was the system adopted for the first stage of the Tahoe lime treatment plant. The integrated type of equipment is sometimes referred to as an upflow clarifier or a sludge blanket clarifier. The latter term can be misleading, however, since not all such units operate with a sludge blanket. A diagram of a unit that does not operate with a sludge blanket is shown in Figure 8-4. Culp and Culp (1) recommend the system of separate tanks because that system allows separate control of each part of the process. They point out that such a system has greater flexibility in points of addition for chemicals. They also point out that in the integrated units with a sludge blanket there may be difficulty in controlling the blanket height and the blanket may become anaerobic, leading to poor phosphorus and solids removal. Pilot studies at other locations have shown that excellent solids removal can be obtained with sludge blanket equipment, but that blanket instability could be a problem. On the other hand, pilot upflow units designed to operate without a sludge blanket have proved to be very effective at Washington, D.C. (2). Results available at this time indicate that both the system of separate tanks and the integrated units without sludge blankets should be considered for design of new plants.

The Tahoe plant is an excellent example of a system with the first stage consisting of separate tanks for each unit process. At the present time the plant must be relied upon heavily as a guide to design of such systems. Description of the equipment and some design parameters are given by Culp and Culp. The rapid mix tank which is located in one corner of the flocculator has about a 30 second residence time at design flow. (The plant usually runs with a daily average flow of about one third of design flow.) Mixing is accomplished with a vertical shaft mixer. The flocculator is a

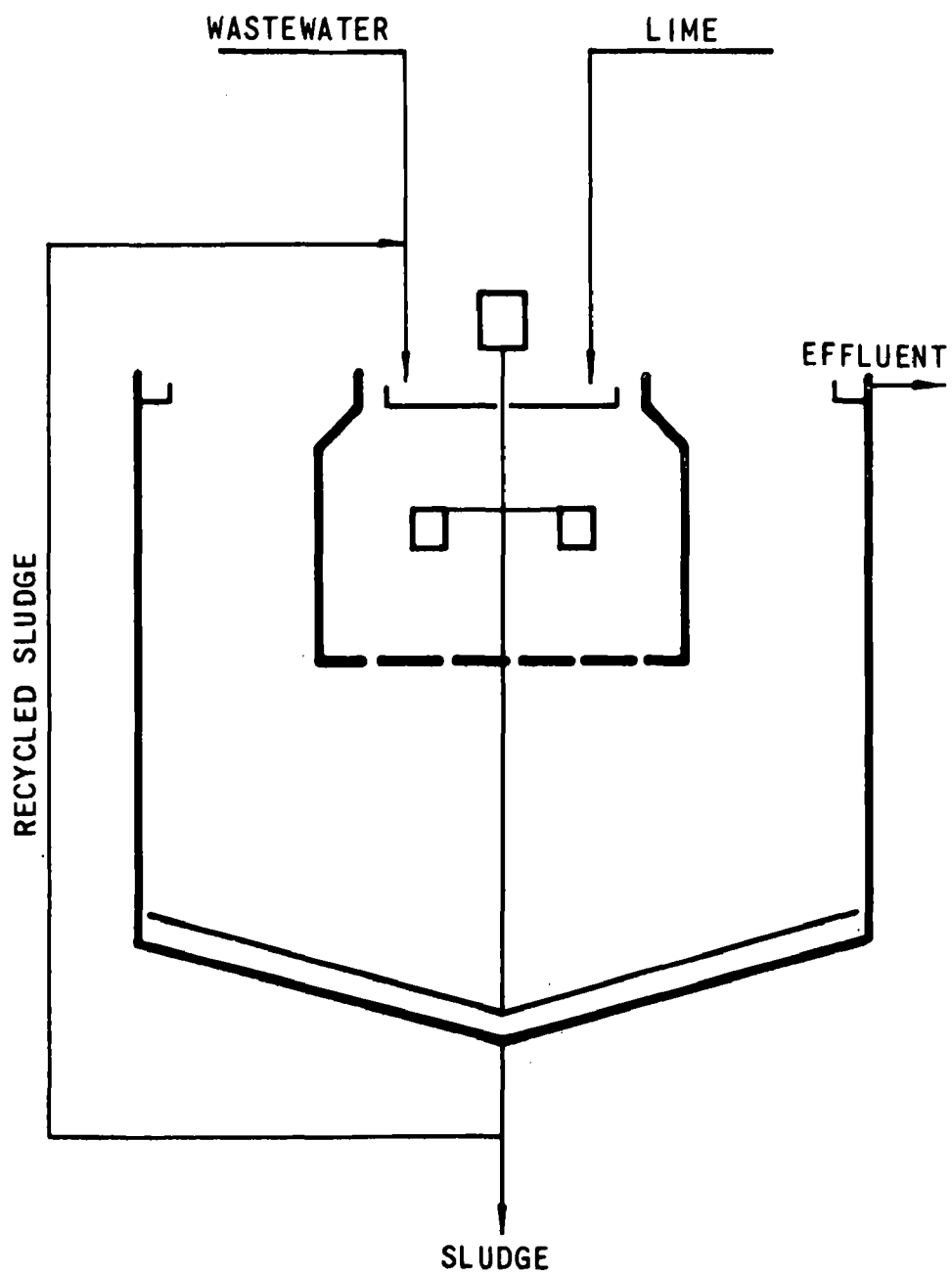


FIGURE 8-4 TYPICAL UPFLOW CLARIFIER

square tank with a depth of 8 ft. At design flow the residence time is 4.5 minutes. The flocculator is provided with air agitation, but operating experience has shown that this agitation is not necessary. The settler is circular with a center inlet. The depth is 10 ft and the design overflow rate is 950 gpd/ft². There are two sludge pumps, one a variable-speed centrifugal, and the other a positive displacement Moyno. Provision has been made to return part of the sludge to the rapid mix tank. The concept of returning sludge to the section of the clarification equipment where precipitation is taking place is called solids contact. The objective is to hasten precipitation and to obtain larger precipitate particles which will settle well.

The ease with which flocculation occurs at Tahoe would suggest that a flocculator for this service presents no design problems. Experience is, however, limited. Pilot testing would be very desirable to determine the flocculating characteristics of a particular wastewater. Jar tests may be of some value. Unfortunately, the effect of solids contact, which experience at Tahoe and elsewhere indicates is beneficial, is difficult to duplicate in a jar test. If jar tests indicated good flocculating characteristics without benefit of solids contact, however, flocculation should be as good or better with solids contact. When pilot studies are carried out, provision should be made for solids contact.

Comments concerning the effect of solids contact in jar tests also apply to settling rates determined from these tests. Rapid settling without benefit of solids contact would be a strong indication that settling with solids contact would also be rapid.

Clarification equipment of the integrated type for use in municipal water treatment is manufactured by a large number of companies. Units of this type used for treatment of wastewater have been essentially of the same design as used for municipal water supplies. The basins are usually circular with the rapid mix and flocculating sections located at the center of the settler. Agitation for mixing and flocculating as well as power for sludge scrapers is provided at the center of the settler. Solids contact is usually provided, and may be carried out with best control by external circulation of sludge to the mixing section. In some designs, however, internal recirculation is used. There is considerable variation in the geometry and complexity of the mixing and flocculating sections depending upon the manufacturer. The unit shown in Figure 8-4 is a relatively simple type. Units from three different manufacturers were used in pilot work at Lebanon, Ohio and Washington, D. C. These and others at additional locations have performed effectively. However, in some cases with operation of a sludge blanket, difficulties have been encountered. A sludge blanket can be eliminated by avoiding designs in which the wall separating the center compartments from the concentric settler reaches close to the bottom of the settler. The design in Figure 8-4 prevents sludge blanket formation.

There has been little effort to modify this type of equipment from use in water treatment to use with wastewater. Experience at Tahoe would suggest, for example, that less flocculation time, and probably less agitation, is required than is generally provided. A unit such as that shown by Figure 8-4 may have about 40 minutes flocculation time at design capacity for water treatment, nearly nine times that at

Tahoe. Whether modifications in such equipment would be worthwhile remains to be seen.

If a designer chooses to use equipment that is on the market, he does not have complete flexibility in specifying the design. Fortunately, equipment satisfactory for water supply treatment has proved satisfactory for wastewater, except that a lower settler overflow rate is necessary. The single-stage treatment system at Lebanon, Ohio was operated at a constant overflow rate of 1,440 gpd/ft². This proved satisfactory even with a sludge blanket. At Washington, D. C., tests were made with settler rates as high as 1,950 gpd/ft² although the average rate was 1,120 gpd/ft². For design, a peak, dry weather overflow rate of 1,200 to 1,400 gpd/ft², with the average somewhat lower, appears reasonable.

8.4.2 Recarbonation

In a two-stage system recarbonation follows the first stage settler. In a single-stage system and following the second stage of a two-stage system, pH reduction by recarbonation may be necessary to make the effluent suitable for filtering or for discharge. An excellent discussion of all aspects of wastewater recarbonation is given by Culp and Culp (1). The reader is referred to that publication, as well as one by Haney and Hamann (5), especially for information on sources of carbon dioxide. Sources include stack gas from sludge incinerators and lime recalciners, CO₂ generators, and commercial liquid CO₂.

The equipment used for contacting the CO₂ and water may be simply a tank with the gas being bubbled through the water. This is the system used at the Tahoe plant where the height of water over the CO₂ source is 8 ft. Pilot studies at Washington, D. C., with a tank depth of 11 ft and a turbine mixer to reduce bubble size and distribute bubbles, showed almost 100% absorption of CO₂.

When considering residence time, the recarbonation tank following the first stage of a two-stage system must be differentiated from the recarbonation tank used just to reduce effluent pH. The residence time of the recarbonator between the first and second stages is not particularly important. At Tahoe the recarbonation tank is only large enough to give a 5 minute residence time at design flow. In the Washington, D. C. studies, the residence time was 15 minutes. A problem that may arise when recarbonating wastewater is foam formation. This is most pronounced when the flow of bubbles is concentrated in parts of the recarbonation tank. The CO₂ distributing system should cover as well as possible the whole lateral area of the tank bottom. This would be especially true if the residence time were as low as 5 minutes.

In the recarbonation tank used just for effluent pH adjustment, sufficient residence time must be allowed for completion of reactions taking place. Culp and Culp recommend 15 minutes. In the Tahoe plant only 4 minutes is provided at design flow, but the water flows to storage ponds with a much longer residence time before the water is filtered. Just as in the case of recarbonation between stages of a two-stage system, good bubble distribution is important for prevention of excessive foaming.

Since one reason for using a two-stage system is to prevent a high calcium concentration in the effluent, the pH to which the water should be recarbonated between stages is that for maximum conversion of the calcium to CaCO_3 . Where lime recovery is practiced, maximum formation of CaCO_3 is also desirable because it results in maximum CaO production. The optimum pH is about 10. At Tahoe, for example, 10.3 was selected because tests showed that pH to give maximum CaCO_3 production.

The pH to be selected when the objective of recarbonation is just pH reduction, will depend on a number of factors. An effluent standard may determine this pH. If stabilizing the water to prevent post precipitation is the objective, the pH may be determined by consideration of the Langelier Index. If, as in the case of Tahoe, a treatment step such as activated carbon adsorption is to follow lime treatment, this will determine pH. In the latter case a pH of about 7.5 has usually been selected.

The CO_2 dose requirements can be calculated from the chemical reactions taking place and a knowledge of the concentrations of the various forms of alkalinity in the water. For design of a new plant, alkalinity data must be obtained from samples of liquor taken from jar tests run at the same pH values as planned for the plant. Results must be considered very approximate, but they do help in sizing CO_2 feeding equipment. Culp and Culp discuss the calculations in detail. It was found in pilot plant work at Washington, D.C. that the CO_2 dose for recarbonation between the stages of a two-stage treatment system can be calculated reasonably well by considering the reduction in calcium concentration that occurs. This reduction is due to CaCO_3 formation with the CO_3^{2-} coming from the CO_2 . The calcium content before recarbonation can be determined satisfactorily from jar tests run at the desired operating pH. The calcium content after recarbonation can also be determined approximately by jar test. Results from several pilot plants indicate the soluble calcium content at the pH of minimum solubility should be about 40 mg/l as Ca. This figure is probably just as good for calculations as a figure obtained from a jar test. The CO_2 dose in mg/l is then equal to:

$$(\text{Ca reduction in mg/l}) \left(\frac{44}{40} \right).$$

A safety factor of about 20% should be added to the calculated dose to compensate for inefficiency in absorption.

8.4.3 Second Stage Flocculation and Sedimentation

In two-stage treatment, recarbonation is followed by the second stage flocculation and settling. Experience with equipment for the second stage of treatment with wastewater is limited. Two very different systems have been tested. The simplest is the system at Tahoe. In the Tahoe plant the equipment used is just a longitudinal settler with a 30 minute detention time and a 2,400 gpd/ft² overflow rate at design flow. No flocculation equipment is provided. Culp and Culp (1) refer to the settler as a reaction and settling basin. It has been found at Tahoe that a significant part of the recarbonation reaction actually occurs in the settler. This is shown by the pH decrease

of about 0.7 unit that occurs between the recarbonation tank and the settler outlet. The clarification that results at Tahoe is unusually good for the simplicity of the equipment.

Jar test and pilot plant work at Washington, D.C. indicated that flocculation after addition of a flocculating aid was required to obtain satisfactory clarification of that water. The pilot tests were run in upflow equipment of the integrated type discussed earlier. Without a flocculating aid, fine CaCO_3 precipitate escaped the settler, even at a settler overflow rate as low as 1,100 gpd/ft². Using 5 mg/l of Fe^{3+} , good operation was observed at overflow rates as high as 1,950 gpd/ft². Hydraulic limitations prevented testing at higher rates. Apparently, the good contact between solids and water during flocculation was beneficial in reducing CaCO_3 supersaturation in the effluent. Precipitation in the filters following the second stage did not occur, and no pH reduction was required.

In this work the sludge from the second stage was returned to the mixing section of the first stage to serve as a weighting agent. In waters of low alkalinity or expected high loads of biological solids, this procedure should be considered.

It is difficult to recommend the amount of flocculation to provide and the overflow rate to use for second-stage treatment. Jar tests would be of little value because of the complication of first raising the pH with lime and then lowering with CO_2 . Pilot testing may be desirable. Such a pilot system would have to include the first stage equipment and recarbonation equipment, although these could be of crude design. In the absence of pilot tests to indicate otherwise, flocculation should be provided in the second-stage system. If filtration is included, a peak dry weather overflow rate of up to 2,000 gpd/ft² in the settler may be used. If filtration is not provided, a somewhat lower overflow rate should be chosen to assure reasonable effluent quality during periods when an upstream part of the system is not operating properly.

8.4.4 Filtration

The last step in a complete lime treatment system is filtration. Although there are a variety of filter types that could be used, essentially all the recent test work has been done with downflow filters of multimedia type. These are essentially the same in design as a rapid sand filter, except that the media are graded from coarse at the filter surface to fine at the filter outlet. This is accomplished by using media of different densities with the largest particles being composed of the least dense material. The result is a filter bed which has far more capacity for removing suspended solids than an ordinary sand filter, without impairing effluent clarity. Both gravity and pressure filters have been used. Each has certain advantages. Generally, however, pressure filters are more appropriate at smaller plants.

Two media combinations have been tested extensively. These are dual-media of coal and sand, and tri-media of coal, sand, and garnet. Filters used at Lebanon, Ohio following a single-stage system (3) and filters at Washington, D.C. following a two-stage system (2) were of the dual-media type. Both contained 18 in. of anthracite

coal over 6 in. of sand. Average particle sizes of the media at Lebanon were 0.75 mm and 0.46 mm; at Washington, D.C., 0.9 mm and 0.45 mm. At 2 gpm/ft² the average run length at Lebanon was about 90 hours using 8 ft of water as the terminating pressure loss. At Washington, D.C., filter run length averaged about 50 hours for an average rate of 3.4 gpm/ft² and a terminating pressure loss of 7 ft of water. High clarity waters were obtained in each case. Backwash in both cases was carried out at 20 gpm/ft². Backwash time was 5 minutes at Lebanon and 10 minutes at Washington, D.C. In the latter case, a surface wash was included.

These run length figures give a rough idea of the results that may be expected at other locations. For maximum run length at a given product quality, the depths of media and their particle sizes should be optimized. Although this was not done for the above-mentioned filters, pressure loss distribution in the filters at Lebanon as reported by Berg, et al. (3) indicates that good solids storage in the anthracite was being obtained. Small pilot filters could be used for optimization of the media.

A reasonable design rate for gravity-flow dual-media filters appears to be about 3 gpm/ft². Higher rates could result in inconveniently short runs during periods of high solids load. Rates much lower than 3 gpm/ft² result in excessive filter costs, not justified by the longer filter runs. Very long filter runs could result in backwash problems from biological activity in the filters. Recommendation of a 3 gpm/ft² rate is based upon having good operation of the second-stage settler. Frequent settler upsets must be avoided.

Tri-media filters are used at the Tahoe plant. These are pressure filters which at design flow operate at 5 gpm/ft². Each bed holds 3 ft of mixed coal, sand, and garnet as supplied by Neptune Microfloc. Two filters are used in series, and are backwashed in series, usually when the head loss reaches 16 ft of water. Run length has varied from 4 hours during heavy solids load to about 60 hours under good conditions. Backwash is carried out at 15 gpm/ft². The backwash water is reprocessed through the lime treatment system. More details are given by Culp and Culp (1).

From the available data, it is difficult to give precise criteria for choosing between dual-media or tri-media filters following lime treatment. The use of garnet allows for a smaller particle size at the bottom of the filter than is possible with sand. In case of floc weakness, the tri-media filter offers, therefore, more protection from turbidity breakthrough. The cost, however, is slightly higher. Where the tri-media fill is not used, filters can still be backwashed at the onset of turbidity breakthrough. Experience with dual-media filters on lime treated water has not shown sudden breakthrough to be an important problem. Where water of the highest clarity is required, tri-media filters may be of most value.

8.4.5 Sludge Handling

The sludge from a lime treatment system may be handled in two general ways. It may be thickened, dewatered, and disposed of or it may be thickened, dewatered, and

recalcined to recover lime for reuse. For small plants, recalcination will not be economically competitive with disposal and should not be considered unless there are restrictions on disposal which make that alternative difficult. It can be assumed generally that for plants over 10 mgd, recalcination of lime sludge will be practical. Recalcination may also be practical at plants somewhat smaller than 10 mgd, depending on the cost of purchased lime and other local conditions.

Data given earlier indicate that the volume of sludge from lime treatment will vary from 1.5% to several percent of feed volume. Sludge concentration will probably be in the range of 1% to 5%. The actual weight of sludge will vary with the chemical composition and amount of suspended solids in the feed water. Values have been observed in the range of 4 to 7 lb/1,000 gal. Sludge production from the first stage of a two-stage system or from a single-stage system can be estimated approximately by weighing the dried sludge from jar tests run at the planned operating pH. Additional sludge produced in the second stage can be assumed to be the CaCO_3 formed from calcium concentration reduction during recarbonation. Calcium content before recarbonation can be obtained from the above-mentioned jar tests and after recarbonation can be assumed to be 40 mg/l as Ca.

Design information for the Tahoe sludge handling system is reported by Culp and Culp. (1) The reader is referred to that publication for further information. The Tahoe plant is the only one which includes recalcination of sludge from the lime treatment of wastewater, and that has been operated for a long period of time. Parameters from that system can be used as a rough guide for design. The gravity thickener has a bottom scraper mechanism and is 8 ft deep. Overflow from the thickener is returned to either the primary clarifier or the first stage of lime treatment. The design solids loading is 200 lb/day/ft² and the design overflow rate is 1,000 gpd/ft². Some preliminary data from pilot work at Washington, D.C. suggest these loadings are high. The Tahoe equipment was sized, however, to handle a volume of sludge equal to about 9% of the plant design flow. The actual volume of sludge should usually be less than one third of that volume. At Washington, D.C., recalcined lime was found to produce a sludge which thickened significantly faster than sludge from virgin lime.

If lime is not to be recovered from the sludge, the underflow from the thickener can be placed directly on drying beds for final dewatering, or it can be dewatered by centrifuge or vacuum filter.

If sludge recalcination is planned, the thickened sludge would be dewatered, by centrifuging or filtering and fed to the calcining furnace. The centrifuge may have advantages over filters for this purpose, such as the ability to separate phosphate sludge from CaCO_3 . Lime from the furnace is stored for reuse. It, along with any new lime, must be slaked before use. Sludge can be pumped from the thickener to the centrifuge. Cake from the centrifuge must be transported by conveyor. At Tahoe the centrate is sent to the primary clarifiers. It has been found at Tahoe that, by operating the centrifuge with less than maximum solids capture, much of the precipitated phosphorus can be retained in the centrate. This results in a higher

quality lime. This mode of operation requires a second centrifuge to remove the phosphorus rich solids from the centrate. These solids may have value for use in fertilizer.

The Tahoe dewatering and recalcining system is described by Culp and Culp. The centrifuge is a 24 by 60 in. concurrent flow type. The calciner is a 14 ft-3 in. diameter, 6 hearth furnace operated at a top temperature of 1,850⁰ F. Other types of furnaces have been used in water treatment plants and these should also be applicable to the sludge from wastewater treatment. The reader is referred to equipment manufacturers for further information about centrifuges and furnaces.

8.4.6 Control of Lime and Carbon Dioxide Feed

The feeding of lime to a lime softening system can be controlled in a number of ways. For the wastewater application, the most appropriate appears to be control of pH, with the pH value being selected for good suspended solids removal or to meet a phosphorus removal requirement. Where flow equalization is employed, pH sensing alone should be sufficient. Where there is substantial diurnal variation in flow, better control can be maintained by flow proportional-pH control. Systems are available for this type of control.

Carbon dioxide dose is also controlled by pH. As in the case of lime feed, flow proportional-pH control would be preferred where there is diurnal variation in flow.

There is a tendency for pH electrodes to become coated with precipitates and lose sensitivity. The precipitate can be dissolved with acid. In some instances, however, it has been found satisfactory to use manual control because of difficulties with the pH control system.

8.4.7 Scale Formation on Equipment

Because the water in a lime treatment system is supersaturated to some degree with CaCO_3 and other precipitating substances, there is a tendency for scale to form on equipment and pipe surfaces. The problem is particularly serious with the lime slurry from the slaker. This can quickly plug the slurry line to the rapid mix tank. There should be easy access to all parts of this line for cleaning. At one small plant a flexible hose was used to feed lime slurry. By periodically flattening or flexing the hose, the scale was removed. The mechanical mixer in the rapid mix tank will also become scaled and must be cleaned.

Scale can also form in sludge lines and the effluent line from the first-stage clarifier. It is recommended that open troughs be used wherever possible. Provision should be made for cleaning lines when open conduits are not possible.

Possible scaling of filters has already been mentioned. Recarbonation before the filters will minimize scale formation.

8.5 Capital and Operating Costs

Because of the short history of lime treatment of wastewater, there is a scarcity of capital and operating cost information. Costs are available for the Tahoe plant and are reported by Culp and Culp (1). Capital cost for the 7.5 mgd lime treating facility was \$1,115,000 and cost of the filters was an additional \$705,000. For 1969 the estimated operating costs exclusive of equipment amortization were 7.3¢/1,000 gal. for lime treatment without filtration and 2.8¢/1,000 gal. for filtration. Amortization of equipment, based on costs adjusted to the 1969 national average, interest of 5% for 25 years, and the assumption of the plant operating at full capacity, would add 2.7¢/1,000 gal. for treatment without filtration and 1.8¢/1,000 gal. for filtration. For the plant operating at full capacity the total cost of operation would then be 10.0¢/1,000 gal. without filtration and 14.6¢/1,000 gal. including filtration. The fraction of the cost resulting from amortization is a substantial 31% even with the assumption of full capacity. There is an obvious need to keep equipment size to a minimum. Flow equalization deserves strong consideration when planning for lime treatment of secondary effluent to minimize the initial cost of new equipment.

Additional costs for lime treatment based on information from Tahoe and other sources have been reported by Smith and McMichael (6). Tables 8-2 and 8-3 show

Table 8-2

CAPITAL COST OF LIME TREATING FACILITIES

<u>Treatment</u>	Cost (\$)		
	<u>Plant Size (mgd)</u>		
	<u>1.0</u>	<u>10</u>	<u>100</u>
Single-Stage without Filtration	100,000	1,200,000	5,500,000
Two-Stage without Filtration	160,000	1,500,000	7,900,000
Dual-Media Filtration	110,000	510,000	2,300,000

Table 8-3

TOTAL COST FOR LIME TREATMENT OF WASTEWATER

<u>Treatment</u>	Cost (¢/1,000 gal.)		
	<u>Plant Size (mgd)</u>		
	<u>1.0</u>	<u>10</u>	<u>100</u>
Single-Stage without Filtration	13	7	4
Two-Stage without Filtration	16	9	6
Dual-Media Filtration	8	3	1.4

costs based largely on their results. Capital costs have been updated to December, 1970. Amortization is at 6% for 25 years. For 1 mgd plants, recalcination equipment is

not included. All lime would be purchased. The applicability of recalcination was discussed earlier. Component costs making up the total cost figures represent an average for the whole country. Local conditions may cause significant deviation from these values.

8.6 References – Chapter 8

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LIME RECOVERY AND REUSE

LIME RECOVERY AND REUSE

by

Robert B. Dean
Chief, Ultimate Disposal Research Program
Advanced Waste Treatment Research Laboratory
National Environmental Research Center
Cincinnati, Ohio 45268

Riverside, California
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LIME RECOVERY AND REUSE ASPECTS

of

ADVANCED WASTEWATER TREATMENT
AS PRACTICED AT SOUTH TAHOE

by

SOUTH TAHOE PUBLIC UTILITY DISTRICT
SOUTH LAKE TAHOE, CALIFORNIA

for the

WATER QUALITY OFFICE
ENVIRONMENTAL PROTECTION AGENCY

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SECTION III

INTRODUCTION

Today there is widespread public recognition of the gross inadequacies of conventional secondary sewage treatment processes in protecting the environment under many circumstances and the need for the practical application of new treatment methods which produce reclaimed water of superior quality.

At Lake Tahoe this need became apparent about ten years ago because of special local conditions which will be described in more detail later in this report. As a result, research and pilot plant studies were undertaken in 1961 to reveal possible new processes for wastewater reclamation. It was found that a tertiary sequence of treatment including conventional activated sludge followed by chemical treatment, mixed-media filtration, and granular carbon adsorption produced remarkable improvements in water quality. These improvements included virtually complete removal of suspended solids, BOD, bacteria, and other substances only partially removed by secondary treatment. In addition, good removals were obtained of COD, color, odor, viruses, phosphates, MBAS, and other substances which are relatively unaffected by secondary treatment.

In 1963, a 2.5 mgd tertiary plant was designed for the South Tahoe Public Utility District incorporating these processes, plus facilities for thermal regeneration of granular activated carbon for flows up to 10 mgd. This plant was built in 1964 and 1965 and was placed into operation during the summer of 1965.

In April 1965, further laboratory, pilot plant, and full-scale plant studies were initiated at Tahoe with WQO, EPA demonstration grant funds. These studies included recovery and reuse of coagulant, nitrogen removal, and data collection in connection with full-scale carbon regeneration and reuse.

In 1966, plans and specifications were prepared to expand the capacity of the entire South Tahoe plant from 2.5 to 7.5 mgd. Facilities were planned for the recovery and reuse of lime as a coagulant and for the incineration of all sludge produced. These plant additions were completed and placed into operation on March 31, 1968. Also an experimental ammonia stripping tower with a nominal capacity of only 3.75 mgd (one half that of the basic plant) was built and placed into operation in November 1968.

This 7.5 mgd Water Reclamation Plant of the South Tahoe Public Utility District is the largest and most complete scale advanced wastewater treatment plant in the world, and was the first to be placed into operation. It has been designated as a National Demonstration Plant by the WQO, EPA under the Clean Waters Restoration Act.

WQO, EPA Demonstration Grant WRPD 52-01-67 not only provides for studies of recovery and reuse of lime as a coagulant and ammonia stripping, but also for complete detailed studies and reporting of the entire plant operation including costs for a three year period ending February 1971.

This report includes: a discussion of the history and purpose of the Tahoe project, a description of the process used, design data for the plant and export system, an outline of the on-the-job training of plant personnel, a description of sample collection techniques and test procedures employed, detailed descriptions of each liquid processing step and each solids handling procedure, detailed results of treatment and plant operation, complete data on actual construction and operating costs, information on the prolonged storage of reclaimed water in Indian Creek Reservoir, conclusions and recommendations, and other miscellaneous related information.

Only the lime recovery aspects are included in this excerpt.

Requirements for Ultimate Disposal of Sludge. Of major importance in process selection are the circumstances related to the handling and ultimate disposal of the sludge produced. In locations where there are available remote, large areas of land, almost any kind of sludge, wet or dry, stable or decomposing, can be, and is, disposed of by hauling or pumping to these land disposal sites. In many places this crude method for sludge disposal probably will not be tolerated indefinitely, but might suffice for the time being. Ocean disposal of sludge has long been an easy way to evade the knotty problems involved in proper sludge disposal. However, the nuisances which have been created and the damages wrought to beaches and coastal waters have aroused the public to the point where this method is now in almost universal public disfavor.

All of the more acceptable methods for sludge disposal involve dewatering of the sludge. The ease with which sludge may be dewatered is a prime factor in unit process selection. There are many alternate ways to process the liquid component of wastewater to secure the desired result at about equal costs, but there are very few ways to satisfactorily

and economically dewater sludge. In sludge from certain wastewaters, dewatering of mixtures of organic-chemical sludges may be satisfactory, but care must be taken to check this out before designing a full scale plant. Favorable pilot plant test results are an important prerequisite.

Another approach which has been used successfully is to keep entirely separate all organic sludge from all chemical sludges. Then, conventional equipment used for either of these types of sludge can be installed. Pilot tests are still highly desirable, even with this approach.

Heat treatment or anaerobic digestion of organic-chemical sludges may condition these mixtures to make possible use of a wide range of dewatering equipment. Chemical conditioners may accomplish this same end, but dosages must be determined to ascertain economic feasibility.

At South Tahoe, the choices in sludge disposal were: incineration, or digestion and hauling the sludge out of the basin provided a disposal site could be found. Because of the long haul involved in sludge export, the decision to incinerate all waste organic and chemical sludges was quite easy as it was favored both by esthetics and economics. Existing sludge digestors for the original 2.5 mgd plant were utilized in the expanded plant for emergency storage of sludge. They also can be used to pretreat sludge for easier dewatering if necessary.

One of the greatest single factors favoring the tertiary sequence of treatment (in which advanced treatment follows conventional biological treatment) is the sludge handling problem. At Tahoe, mixtures of primary, waste activated, and chemical sludges proved to be prohibitively expensive to dewater, but organic sludges and chemical sludges which were kept separate dewatered readily and at low cost. This is one of the most important factors to be evaluated prior to final plant design, particularly in the decision between physical-chemical treatment and the tertiary sequence.

SECTION VI

THE TAHOE PROCESS FOR WASTEWATER

RECLAMATION

Water Reclamation Plant. The South Tahoe water reclamation plant is the most advanced full-scale wastewater treatment plant in the world, although other similar plants are now under construction in other places.

Treatment of wastewater consists of two basic parts, liquid processing and solids handling. The first two steps of liquid processing are the conventional ones of primary, or solids separation, and secondary, or biological oxidation. In addition, the advanced treatment provides chemical treatment and phosphate removal, nitrogen removal, mixed-media filtration, activated carbon adsorption, and disinfection.

The solids handling system provides for incineration of biological sludge, regeneration and reuse of granular activated carbon, and recalcining and reuse of lime, all by means of multiple hearth furnaces. The furnaces are equipped with scrubbers and after-burners to prevent air pollution.

A detailed description of the treatment plant processes will be given in the normal order which they occur in flow through the plant, first for the liquid processing and then for the solids handling.

Figure 5 is a schematic flow and process diagram.

Liquid Processing. All wastewater is pumped to the reclamation plant. Plant influent may be prechlorinated for odor control. The raw wastewater is then passed through a barminutor which screens and shreds the coarse solids, then through Parshall flumes which measure the plant inflow.

The water then passes to either or both of two primary settling tanks where the liquids and solids are separated by sedimentation. In addition to the raw wastewater, the primary tanks may also receive the overflow from the lime mud thickener, and the centrate from the lime centrifuge or the sludge centrifuge. If desired, lime or polymer may be

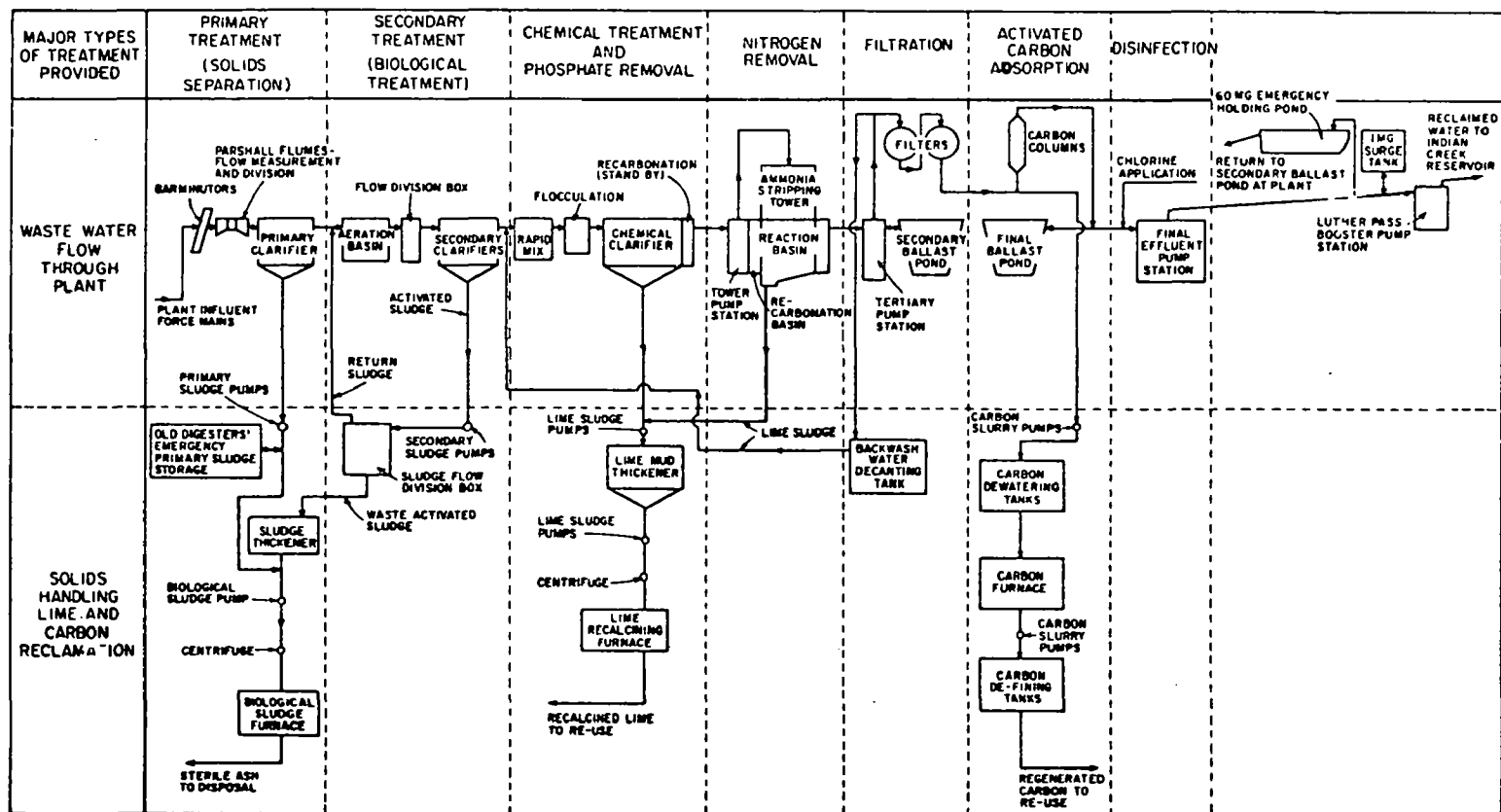


FIGURE 5
SCHEMATIC FLOW AND PROCESS DIAGRAM

added ahead of the primary tanks. One of the primary tanks is a rectangular basin, and the other is a circular basin. Both are equipped with mechanical collectors for continuous sludge removal.

The next step in the process is secondary treatment or biological oxidation. One-third of the plant capacity utilizes the conventional plug-flow activated sludge process with diffused air. The other two-thirds of the secondary treatment capacity is provided by a complete mix system using a combination of diffused air and mechanical mixing, or surface aeration. Secondary settling takes place in either of two circular basins, each of which has its own sludge recirculation pump station. Chlorine may be added to the secondary settling basin influent for control of sludge bulking as necessary.

Secondary settling is the end of the processing in conventional treatment plants, so that the additional treatment provided beyond this point is all in the nature of advanced waste treatment.

The secondary effluent together with waste filter backwash water enters a rapid mix basin where violent mechanical stirring occurs, and where lime is added to pH=11+, which corresponds to a dosage of about 400 mg/l of lime. Next comes a period of slow mixing and flocculation by air agitation. Following this, a polymer is added, usually in the amount of about 0.1 mg/l, as the flocculated high-pH water flows to the chemical clarifier. The chemical clarifier, a circular basin, separates the liquid from the rather large quantities of lime sludge.

Overflow water from the chemical clarifiers flows by gravity to a sump where either of two pumps lifts it to the top of the nitrogen removal tower, the first tower of its kind used in a municipal wastewater plant. The tower has a nominal capacity of 3.75 mgd (one-half plant capacity), and is the only part of the full-scale plant which is still considered to be experimental in nature. Water pumped to the tower has pH=11+, indicating that the ammonia is virtually all present as dissolved gas, rather than as ammonium ion in solution. Large quantities of air must be circulated through the tower for best efficiency, so that an open packing must be used in the tower to minimize head losses and power requirements for the circulating fan.

The water is distributed uniformly from a horizontal tray across the top of the packing, which is made of treated hemlock slats spaced at 1.5 in. vertically and 2 in. horizontally. As the water strikes a slat, droplets are formed. Surface film thickness in the droplets is at a minimum as they are forming, and this favors the escape of ammonia gas from the droplet. By circulating a great amount of air through the tower, the

air surrounding the droplet is kept at a low ammonia concentration, promoting a maximum transfer of ammonia from the water to the air. Once droplets are fully formed, very little further transfer of ammonia takes place, so the droplets are coalesced on top of the next slat below and new droplets are formed as the water falls off the slat, allowing further escape of the ammonia. This process is repeated about 240 times in one pass through the 24-ft-high tower.

Air in the tower flows across the descending droplets. The air enters through the side louvers and travels horizontally to a central plenum, where it is discharged vertically upward through a fan that has a maximum capacity of 700,000 cfm. Tower loading rates are about 2.9 gpm of water per square foot, and about 390 cfm of air per gallon of wastewater. The efficiency of the tower in removing ammonia will vary from 30 to 98 percent, depending principally upon air and water temperatures and to a lesser extent upon hydraulic loading and air supply.

From the catch basin beneath the nitrogen removal tower, the tower effluent passes over a weir for flow measurement, and then into another basin beneath the tower, which contains three sections. The first section is the primary recarbonation basin. Here, carbon dioxide gas, supplied by compressing stack gasses in the incinerator building, is dissolved in the water to lower the pH from 11 to 9.6. The recarbonated water is then held in a contact basin for 30 minutes or more to allow complete formation and some settlement of calcium carbonate. This basin is equipped with equipment for continuous sludge removal. The third section of the basin is the secondary recarbonation chamber, where the pH is reduced from 9.6 to any desired level, usually 7.5, by further addition of carbon dioxide.

The recarbonated water then flows through two ballast ponds in series. The ballast ponds are used to store water for backwashing the mixed-media filters, and to reduce peak flows to the filters and carbon columns.

Water is pumped from the ballast ponds to the filters and carbon columns. Ordinarily about 5 mg/l of alum are added to the filter influent water. Three pairs of mixed-media filter beds, developed especially for filtering waste water, handle the plant flow.

The mixed-media bed, with pores graded coarse to fine in the direction of flow, is composed of coarse coal, with a specific gravity of 1.4; medium-sized sand, with a specific gravity of 2.65; and fine garnet, with a specific gravity of 4.5. A properly graded bed made of these materials provides an almost ideal coarse-to-fine filter. Because of the different specific gravities of the three materials, the particles retain their desired position in the depth of the bed during backwashing. The

fine filter media is supported on a conventional bed of graded gravel, to which has been added an important new feature, a 3-in. layer of 16-mesh garnet. This extremely heavy material (specific 4.5) positively prevents any movement of the gravel bed below, and also prevents any penetration of the fine garnet above into the gravel supporting bed. The filters are equipped with rotary surface washers, and rotation indicator lights. Good surface wash is essential to proper operation of the coarse-to-fine filters because of the large quantities of particulates removed from the water and stored throughout the depth of the bed.

Each pair of beds comprises a filter unit, and they operate in series both during filtration and backwash. A total depth of 6 ft of fine media is provided in each pair of beds. With coarse-to-fine filters, the length of filter run is almost directly proportional to the depth of fine media, so that the length of run for the two beds in series is about double that for a single bed. By backwashing the two beds in series, about half as much backwash water is discharged to the decant tank and then returned at a slow rate to the rapid-mix basin for reprocessing.

When a pair of filters is plugged or no longer produces a high quality effluent, it is automatically taken off the line, backwashed, filtered-to-waste, and restored to service. This is all done automatically and monitored through a control panel located in the filter building.

The filtered water next flows under pressure to the eight carbon columns, which operate in parallel. Each column is 12 ft in diameter by 24 ft high, and contains about 22 tons of 8 X 30 mesh granular activated carbon. Flow in the columns is of the moving bed, countercurrent type - that is, water flows from the bottom to top of the column, while movement of the carbon is down, the fresh carbon being added at the top and the spent carbon removed at the bottom. This system permits frequent removal and corresponding addition of carbon for maximum operating efficiency.

The water is in contact with the carbon for a period of 15 to 25 minutes, during which the adsorption of organics by the carbon takes place. The carbon column effluent is colorless, odorless, low in organics, and sparkling clear.

The high quality of the water following complete treatment vastly improves the efficiency of chlorination, the final step in the liquid processing. As compared to chlorination of ordinary secondary effluent, the chlorination of the reclaimed water at South Tahoe is many times more effective, as virtually all of the chlorine-demanding materials, except ammonia, present in the secondary effluent have been removed. With rapid violent mixing at the point of chlorine application, good disinfection can

be accomplished in the presence of ammonia. One explanation offered is that chlorine reacts more rapidly with bacteria and viruses than with ammonia.

Solids Handling System. As mentioned previously, all solid plant wastes are processed in multiple hearth furnaces. The biological and waste chemical sludges are incinerated, the lime mud is recalcined and reused in the process, and the spent granular carbon is regenerated and reused.

The materials to be incinerated include primary and waste activated sludge, waste chemical sludge, screenings, skimmings, and the centrates from the lime and sludge centrifuges.

A mixture of all of these materials is pumped to the sludge centrifuge, which operates at about 1,600 rpm, where a polymer is added and the sludge is partially dewatered to a solids content of about 19 percent in the cake. The cake is conveyed by belt to a multiple hearth furnace; the fuel is natural gas. The furnace is operated at about 1,600 °F and reduces the sludge to an insoluble sterile ash, which may be disposed of on the plant grounds. The furnace stack gas is cooled to 110°F and scrubbed. There is no odor, smoke or steam plume, and the discharges meet all air pollution codes. When the sludge dewatering or incineration equipment is out of service, the old sludge digesters may be used for sludge storage.

Handling of the spent lime mud is similar to that just described for biological sludge. Lime sludge from the chemical clarifier is pumped to a sludge thickener which thickens the sludge to about 8 percent solids. The thickened sludge is then pumped to a centrifuge for further dewatering at 1,600 rpm. The cake contains about 40 percent solids. The lime is recalcined at about 1,850°F to a calcium oxide content of about 50 to 80 percent. Again the stack gas is cooled and scrubbed so that there is no air pollution. Only about 75 percent of the lime sludge is recalcined and reused. The other 25 percent is wasted, mostly hydroxyapatite, in the centrate to the primary clarifier and thence to the sludge furnace, or directly to the sludge furnace. Alternately, the centrate from the lime centrifuge can be dewatered in a second centrifuge and the cake incinerated in the sludge furnace.

The recalcined lime is conveyed pneumatically to a storage bin, and then reused in the process. Makeup lime is unloaded from trucks pneumatically and stored in a separate bin. Separate gravimetric feeders and slakers are provided for the recalcined and makeup lime. Many of the furnace operations are controlled from a panel on the main floor of the incineration building.

The carbon regeneration equipment is located in the filter building. After the carbon becomes saturated with materials removed from the wastewater, it loses its capacity to absorb certain organics, and must be regenerated. Originally in 1965, a break-through of MBAS (or detergents) was the indicator that regeneration was needed. With the advent of the soft detergents, a high COD content in the carbon column effluent has become the indicator. In regeneration a carbon column is pressurized and the carbon slurry is drawn off the bottom of the column to one of the dewatering bins. Here the free moisture drains off in about 10 minutes, leaving carbon with a moisture content of about 40 percent, which is suitable for introduction of carbon to the furnace. Spent carbon is fed to the furnace at a controlled rate by a screw conveyor equipped with variable speed drive. The carbon regeneration furnace is operated at about 1,700°F in a limited oxygen atmosphere with the addition of steam. The rate of feed to the furnace and the hearth temperatures are controlled by the apparent density of the regenerated carbon, which is held at 0.48 to 0.49. As a check on regeneration efficiency, iodine numbers (a relative measure of adsorptive capacity) are run on carbon samples in the laboratory. The carbon is regenerated to full virgin activity with an attrition loss that is about 8 percent per cycle. The regenerated carbon is cooled in a quench tank, pumped by a diaphragm slurry pump to wash tanks, washed to remove carbon fines, and then returned to the top of the carbon column.

SECTION VII

PLANT DESIGN DATA

The principal design criteria for the South Tahoe Water Reclamation Plant are tabulated below.

Item	Amount	
Plant design average flow	7.5	mgd
Peak flow rate (except as noted below)	15.0	mgd
Peak flow rate (filters and carbon columns)	8.2	mgd
Maximum hydraulic rate	20.0	mgd
Plant design BOD (summer)	325	mg/l
Plant design BOD (winter)	250	mg/l
Plant suspended solids (summer)	200	mg/l
Plant suspended solids (winter)	150	mg/l
Water temperature (summer)	17°	C
Atmospheric pressure (elevation 6,300 ft)	11.6	psi
Primary clarifier No. 1		
Surface area	2,350	sf
Flow	2.7	mgd
Overflow rate	1,150	gpd/sf
Primary clarifier No. 2		
Surface area	7,850	sf
Flow	4.8	mgd
Overflow rate	610	gpd/sf
Aeration basins 1,2, and 3, plug flow		
Flow	2.7	mgd
Volume	115,000	cf
Detention (without recycle)	7.5	hrs
BOD loading	50 lbs/1,000	cf

Item	Amount	
Pumps to tertiary plant		
No. 1	1,900	gpm
No. 2	3,800	gpm
No. 3	4,200	gpm
Surface wash booster	500	gpm
Mixed media filters		
Flow	8.2	mgd
Units, 3 sets of 2 series beds		
Hydraulic loading	5	gpm/sf
Backwash rate	15	gpm/sf
Area each bed	380	sf
Surface wash flow	0.6	gpm/sf
Waste backwash water receiving tank		
Capacity	80,000	gals
Carbon columns (8), upflow countercurrent		
Flow	8.2	mgd
Carbon volume, each column	1,810	cf
Carbon depth, effective	14	ft
Contact time	17	min
Hydraulic loading	6.5	gpm/sf
Chlorination equipment		
Three feeders, each	2,000	lbs/day
Carbon regeneration furnace, 6-hearth, 54-inch diameter, gas-fired		
Capacity, dry carbon	6,000	lbs/day
Sludge dewatering equipment, concurrent flow centrifuges, 24" x 60"		
Organic sludge		
Number	2	
Capacity, each, dry solids	450	lbs/hr
Lime sludge		
Number	1	
Capacity	1,650	lbs/hr

Item	Amount
Sludge incineration furnace, 6-hearth 14'-3" diameter, gas-fired Capacity, dry solids	900 lbs/hr
Lime recalcining furnace, 6-hearth 14'-3" diameter, gas-fired Capacity, dry CaO	10 tons/day

SECTION XVII

LIME RECOVERY AND REUSE

General. At a flow of 7.5 mgd through the water reclamation plant approximately 17 tons (dry CaO basis) per day of lime mud would have to be dewatered and disposed of. Since about 93% by weight of this lime mud is in the form of calcium carbonate, disposal costs would include not only dewatering and disposing of about 34 tons of water and solids but also the loss of recoverable calcium oxide. By recovering the lime through recalcination, the total blow-down of waste solids is reduced to about 1.5 tons of dry solids. The cost of recalcined lime as shown later would be slightly more than that of new lime at 7.5 mgd; however, at this flow the reuse of lime reduces by a factor of 20 the amount of water and sludge to be disposed of and, therefore, effects a substantial cost savings.

Physical System. Lime mud is pumped from the chemical clarifier and recarbonation reaction basin to a 30-foot diameter gravity thickener, with a design overflow rate of 1000 gal/ft²/day. Thickened lime mud is pumped to a 24" x 60" solid bowl concurrent flow centrifuge. The cake from the centrifuge is carried by a belt conveyor to a 14.3 foot diameter, six hearth furnace in which calcium oxide and carbon dioxide are produced. The recalcined lime is conveyed out of the furnace by gravity through a crusher to a thermal disc cooler where lime temperatures are lowered from 700°F to 100°-150°F, and then into a rotary air lock. The recalcined lime is pneumatically conveyed from the rotary air lock to a 35-ton capacity recalcined lime storage bin for eventual reuse. Stack gases, rich in carbon dioxide, are scrubbed in a multiple tray scrubber before being exhausted to the atmosphere. A portion of the gases are recycled to the recarbonation system. See Figure 41.

Solids wasting must be performed continuously to maintain an acceptable calcium oxide content in the recalcined lime. Wasting can be accomplished by feeding recovered lime to the primary clarifier, by diverting part of thickener influent to the primary clarifier, by conveying lime mud cake from the centrifuge directly to the organic sludge furnace or by using the centrifuge to classify the phosphate and other inerts into the centrate and the calcium carbonate into the dewatered cake conveyed to the

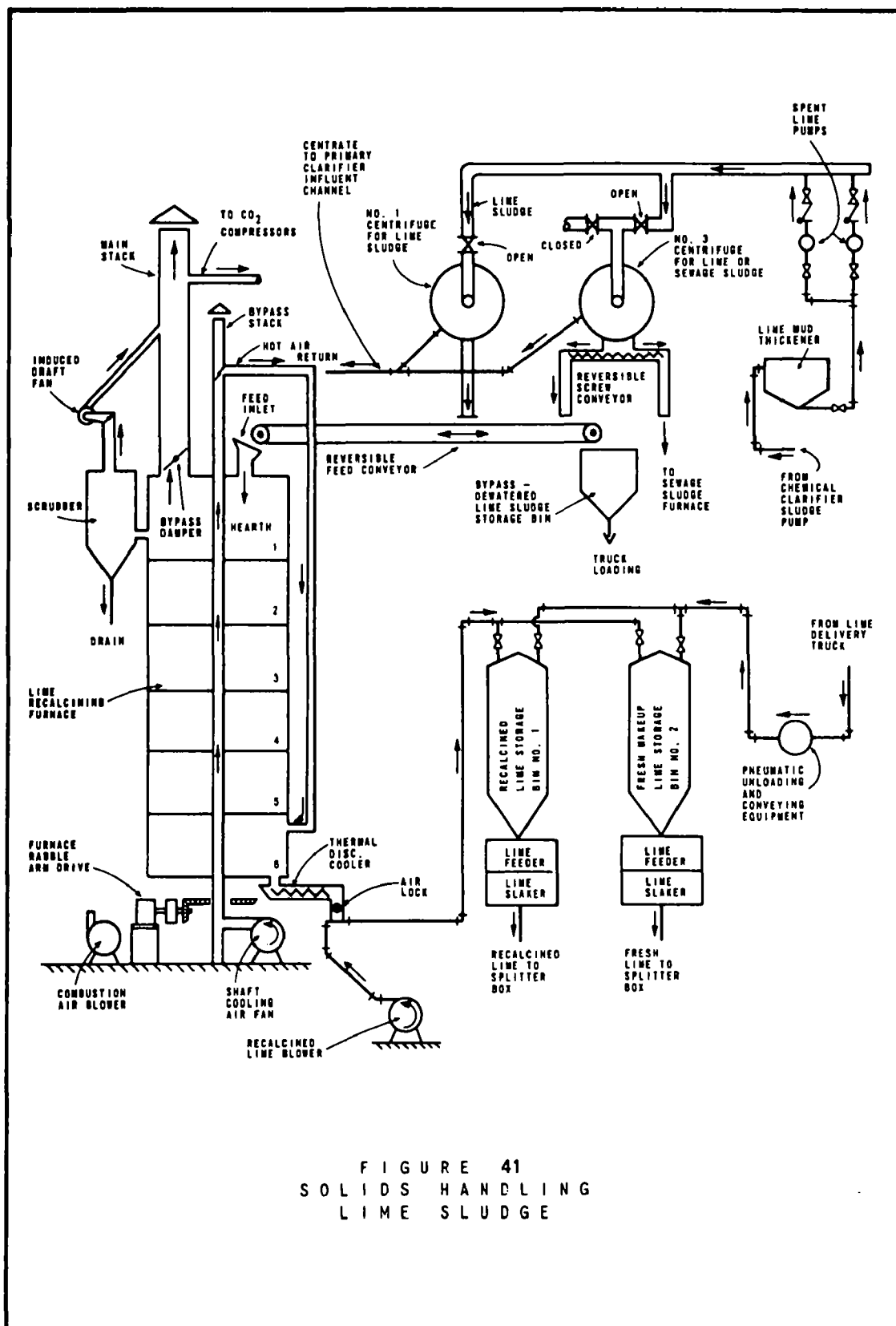


FIGURE 41
SOLIDS HANDLING
LYME SLUDGE

recalciner. A second centrifuge dewateres the centrate from the first machine and its cake is conveyed to the organic sludge furnace.

The centrate from the lead centrifuge may be returned to the primary clarifier instead of directed to the second machine. The second centrifuge's centrate is returned to the primary. The spillage from the lime conveyor belt is collected in a tray and returned to either the primary clarifier or to the lime mud thickener.

Operating Practice. Since plant startup in 1968, steps have been taken to remove all active lime or waste lime mud streams from the primary clarifier. As described in the "Chemical Treatment" section, feeding active lime to the primary clarifier led to organic sludge dewatering problems and to some extent adversely affected primary clarification. Waste lime mud streams being returned to the primary clarifier also affected, but to a lesser extent, the dewatering characteristics of the raw and waste activated sludges.

In January, April and May 1970, and from July 1970 to date, the lime centrifuge has been used to classify inert materials out of the feed to the recalcining furnace. At the same time, the classified centrate from the first centrifuge has been dewatered and clarified in a second centrifuge and dried in the organic incineration furnace. The degree of classification and resulting economies are discussed later in this section.

Lime Mud Thickening. Total solids tests were used to evaluate the efficiency of the lime mud thickener. To determine the amount of calcium lost over the thickener launder the following procedures were used. Basin overflows and sludge withdrawals were measured to determine the influent flow. The amount of calcium in the thickener influent was determined by heating the influent total solids samples in the laboratory muffle furnace at 1850°F for two hours and then determining the available CaO content. By this method, the calcium hydroxyapatite, $\text{Ca}_5\text{OH}(\text{PO}_4)_3$, was not included, and the value reflected only the calcium in CaCO_3 and $\text{Ca}(\text{OH})_2$. Part of the influent sample was filtered 0.45 u millipore filter and calcium hardness was run on the filtrate to determine the $\text{Ca}(\text{OH})_2$. The difference in the two provided the pounds of usable calcium available for thickening.

The pounds of usable calcium in the thickener overflow were found by determining the difference in calcium content between acidified and filtered samples.

The amount of usable calcium lost over the thickener launder was evaluated over an eight hour composite sampling period. Using the procedures described above, 0.04% by weight of the usable calcium coming into the thickener was lost over the weir.

The true ability of the thickener to concentrate lime mud solids could not be evaluated due to low plant flows. Thickener influent percent solids were about 1% by weight. At chemical clarifier flows of 2.8 mgd and thickener underflow rates of 36 gpm, the lime mud was thickened to 4.9% solids; whereas at 3.7 mgd through the chemical clarifier and 15 gpm thickener underflow rate, lime mud was thickened to 8.3% solids.

Lime Mud Classification and Dewatering. During the lime centrifuge acceptance tests, three bowl speeds were evaluated to determine the optimum centrifugal bowl speed. The three bowl speeds evaluated were 1600, 1800, and 2200 rpm. Each of the three evaluations were conducted at 140:1 gear reduction ratio. The function of the gear unit is to drive the conveyor at a fixed speed relative to the bowl. For the three bowl speeds above, the conveyor revolved at approximately 11.5, 13, and 15.5 rpm. For all three bowl speeds evaluated, the cake solids ranged from 36-38%, and recovery or capture remained fairly constant. As a result, the lowest bowl speed of 1600 rpm was chosen for plant operations because of the reduced wear and maintenance problems.

Chemical Addition. Since the lime centrifuge is used primarily for classification purposes and not dewatering, polyelectrolytes were not used to condition the thickened mud prior to centrifuging. In the next section the excellent classification results are discussed. The addition of a polyelectrolyte to the centrifuge feed would defeat these purposes.

Classification Evaluation. Two methods were used to evaluate the centrifugal classification of the inert calcium hydroxyapatite out of the lime solids recovery stream. The first was to make four, ten-minute sampling runs with high and low feed rates and pool depths within the centrifuge to optimize the classification process. An approximate twenty minute interval was allowed after making adjustments to the centrifuge to insure steady state conditions for a particular set of variables. The centrifuge bowl speed was kept constant at 1600 rpm. During each ten-minute period a sample was composited uniformly on the feed, centrate, and cake streams. The centrifuge pool depth, cake volumetric flow rate and wet

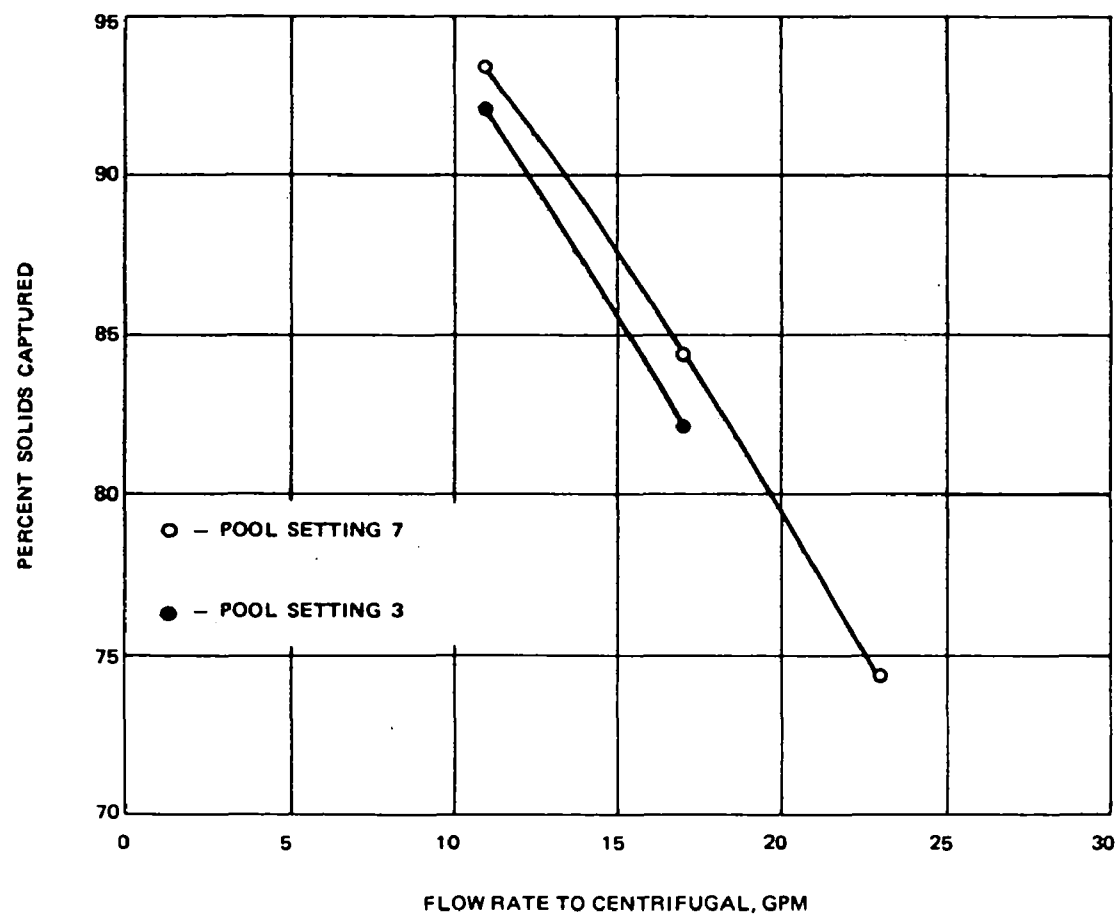
density, and centrate flow rate were recorded. The composited samples were analyzed for total solids and then calcined in a muffle furnace at 1750°F. The total solids, percent capture and centrate flow rate were used to compute the feed rate. After calcination, the samples were analyzed for CaO, and samples were dissolved in acid and analyzed for magnesium and orthophosphates.

With the volumetric flow rate and phosphate and calcium results, balances were made around the centrifuge for usable calcium, and the inert materials, magnesium and phosphate.

The second method was to composite a sample of centrifuge feed, centrate, and cake every two hours for an eight hour period. The centrifuge centrate flow rate was recorded every two hours also. As in the previous evaluation, the centrifuge bowl speed was maintained at 1600 rpm. The resulting composite samples were analyzed for total solids, calcined at 1750°F for 2 hours, and then the usable calcium was determined by analyzing for percent calcium oxide. The dilutions and analysis for phosphate used for the first evaluation were repeated. The percent capture for solids was computed with the analysis information, and the appropriate balances around the centrifuge for usable calcium and phosphate were determined.

Classification Evaluation Results. A brief explanation of the effects of changing the pool depth in a centrifuge is necessary before the classification results are explained. If the pool depth is increased, the cake should get progressively wetter. It is possible to increase the pool depth to the extent that the cake leaves the centrifuge as a wet slurry. However, if the pool depth is decreased to just below this point, the maximum capture should be attained. Conversely, if the pool depth is decreased, the centrate should have a higher solids content, lower percent solids capture, but the cake should be drier. The centrifuge pool depth setting is on a scale of 1-10, with a low number corresponding to a low pool depth and a high number, a high pool depth. As Figure 43 indicates, the centrifuge performed as was expected, the higher pool depth produced the higher captures for similar flow rates. The flow rate to the machine also has a significant effect on the solids capture, with a marked decrease in solids recovery at the higher flows. The highest flow rate, 23 gpm, had a feed solids content of 6%, whereas the other four data points had feed solids contents of 8%. Also the highest flow rate data point was derived from an eight-hour sampling period, and the other four points were from ten minute sampling periods.

FIGURE 43
PERCENT SOLIDS CAPTURED VS FLOW RATE TO CENTRIFUGAL



The removal of inert materials by classification into the centrate stream provides many benefits to the lime treatment and solids recovery system. A significant savings in fuel usage is realized, which is described later in the recalcining portion of this section. Also the size of the waste stream is reduced, a smaller amount of usable lime is lost, which results in lower makeup lime dosage and costs. The effect of solids capture or recovery in the centrifuge on the removal of inert materials from the feed to the recalcination furnace by classification is shown in Figures 44 through 47. The data in Figures 44 and 45 is derived from the two previous sampling runs described and from three runs of the optimization of the recalcination furnace described later. As would be expected, higher captures of solids from the centrifugal feed will result in a lower percent increase in the usable calcium or active lime, CaO , in the cake. At lower captures the lower specific gravity inert materials can be separated from the usable lime. This is very evident in Figure 44, at 90% solids capture there is only a 5% improvement in the percent CaO of cake over the feed to the centrifugal. At 70-75% solids capture, a 12-15% improvement in the cake percent CaO over the centrifugal feed can be expected. Correspondingly in Figure 45, at 95% capture practically all of the usable calcium is conveyed to the recalcination furnace, but so are a majority of the inert materials. At lower solids recoveries a portion of the usable calcium is lost in the centrate, but the higher active lime or CaO content of the cake from better classification of the inerts more than offsets this loss.

Since one of the high priorities of the treatment at South Tahoe is to remove phosphorus from the wastewater, the logical question is how efficient is the phosphorus removal from the lime solids recovery stream. In Figure 46, the effect is shown of solids capture and classification on the amount of phosphate wasted in the centrate. The four data points above 80% capture are taken from the ten-minute sampling periods and the other point is taken from the eight-hour period. At 90% capture or 10% of the solids entering the centrifuge being removed in the centrate, a 20% reduction in phosphate in the cake can be expected. A centrate containing 20% of the solids entering the centrifuge will have almost 40% of the phosphate entering the centrifugal. According to the trend indicated in Figure 46, 90% of the phosphates in the centrifuge feed can be wasted to the centrate at 75% solids capture or recovery, which corresponds well with Figure 44, indicating a near maximum increase in the cake CaO over the centrifuge feed CaO around 75% capture of solids. The removal of magnesium from the centrifuge feed to the centrate by classification is shown in Figure 47. The four data points shown in Figure 47, are from the ten-minute sampling runs; magnesium was not analyzed in the eight-hour run. The fourth point in Figure 47 was disregarded in drawing the

FIGURE 44

EFFECT OF CENTRIFUGAL CAPTURE ON THE CHANGE IN
CALCIUM OXIDE CONTENT OF THE CENTRIFUGAL CAKE

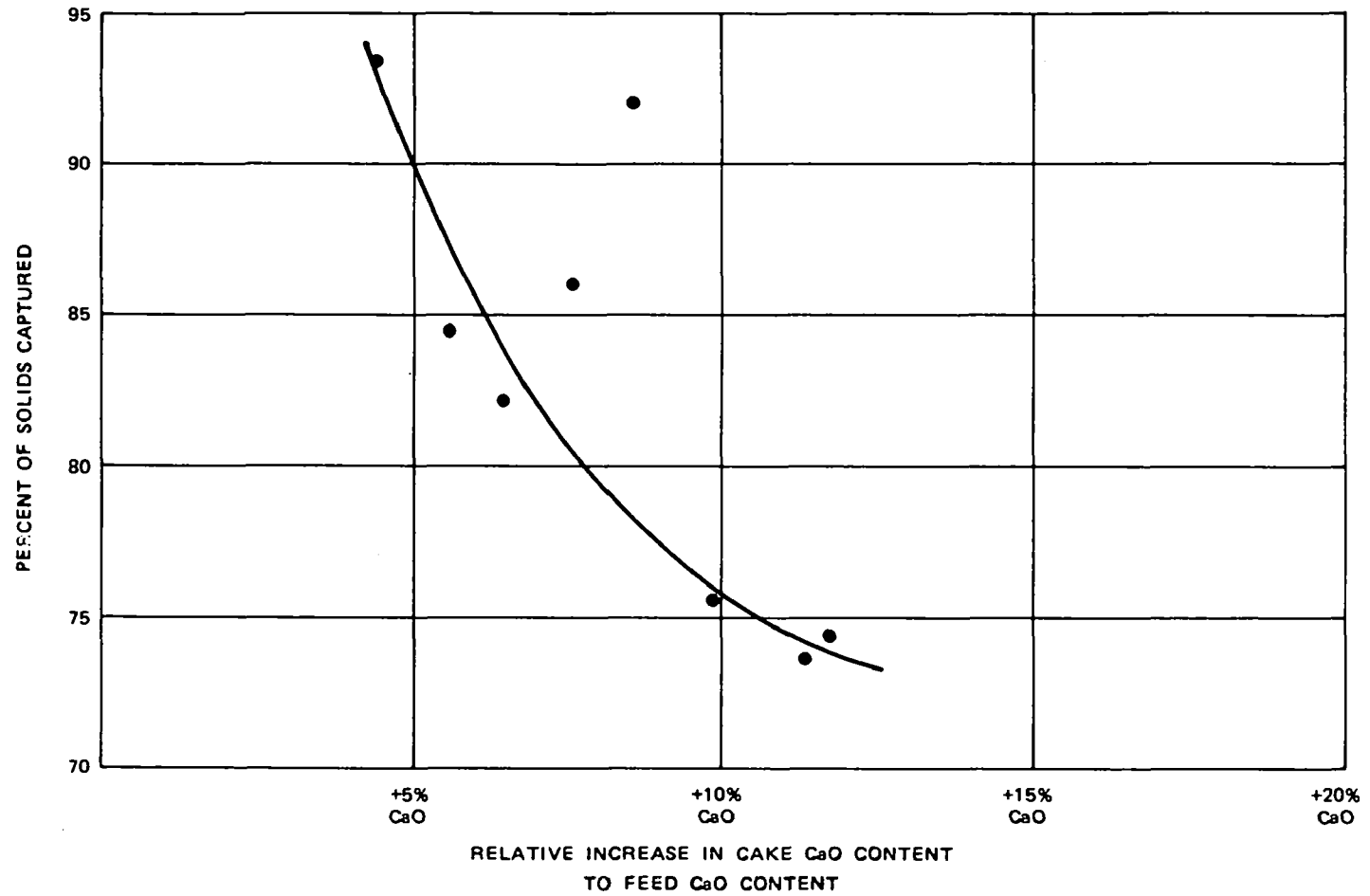


FIGURE 45
PERCENT SOLIDS CAPTURED
VS
PERCENT OF USABLE CALCIUM IN FEED CONVEYED TO FURNACE

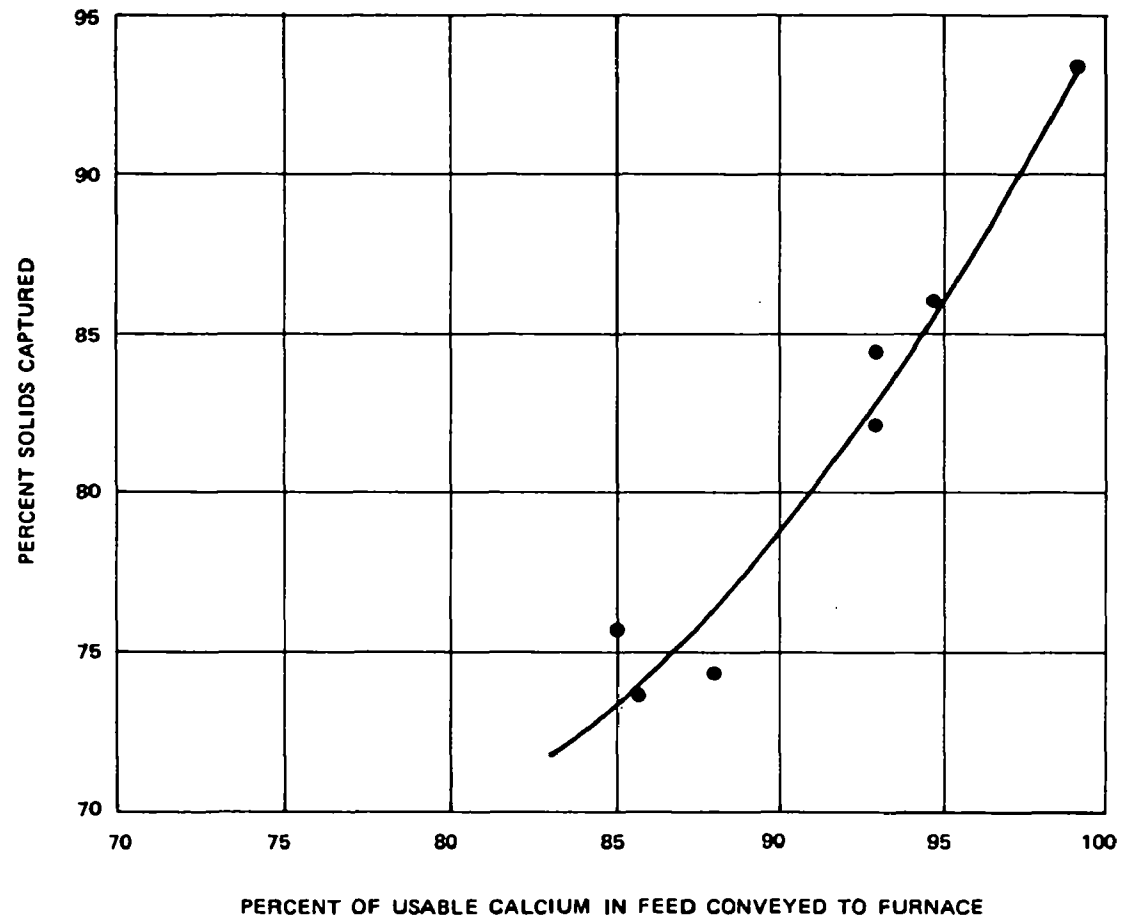


FIGURE 46
PERCENT SOLIDS CAPTURED
VS
PERCENT OF PO_4 IN FEED WASTED IN CENTRATE

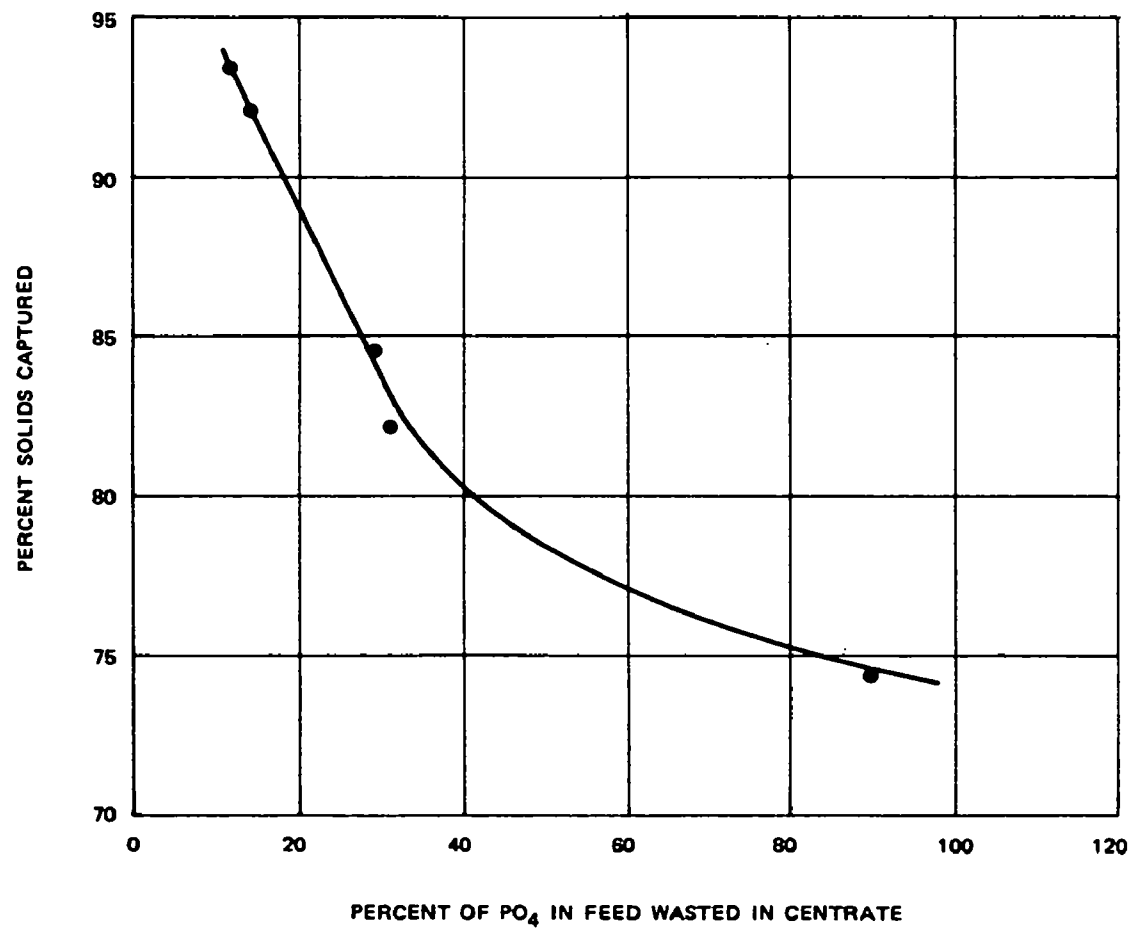
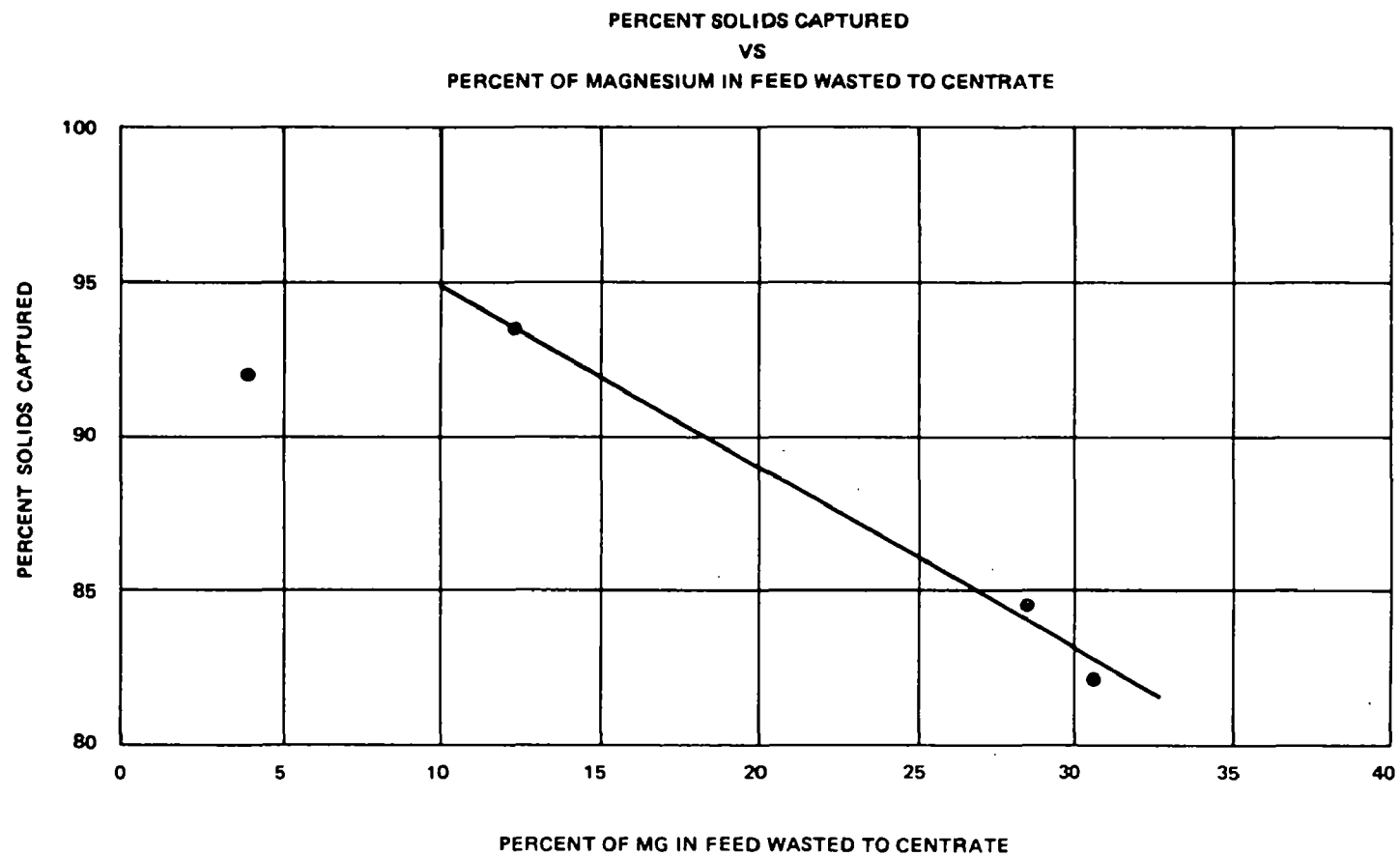


FIGURE 47



line, since the analysis for this run resulted in a 15% error in the magnesium balance around the centrifugal. However, the same trend as the wasting of phosphates in Figure 46 is indicated. At 80% capture, approximately 35% of the magnesium entering the centrifuge is classified into the centrate.

To further exemplify the classification abilities of the lime centrifuge the organic sludge and recalcination furnaces' scrubber waters were analyzed for ortho phosphorus (mg/l PO₄-P) for three different periods of operation. The first period was under normal classifying operations with the lead lime centrifuge classifying and the swing centrifuge dewatering the centrate and conveying the waste inert materials to the organic sludge furnace. The second period was when the recalcination furnace was removed from service for a short period for maintenance, but the organic solids furnace continued to operate. The third period was when the lead lime centrifugal was not used for classifying, but dewatering only, and the resulting cake was being recalcined. The average results of the three periods are shown in Table 14.

TABLE 14

SCRUBBER WATER

ORTHO PHOSPHORUS CONTENT, mg/l PO₄-P

Operation Mode	Sludge Furnace		Recalcination Furnace	
	Total	Soluble*	Total	Soluble*
Classification	17.95	.56	2.4	.32
Not Recalcining	1.54	0.08	0.19	0.07
Lime Centrifuge for dewatering only	3.59	--	12.02	--

*Soluble defined as passing a .45 u filter.

The water reclamation plant effluent is used for the source of supply for the furnace scrubbers. The normal ortho phosphorus content of the plant effluent is in the range of .06 to 0.1 mg/l. When the recalcination furnace was out of service, it's scrubber water phosphorus content was barely affected. During the same period, when the sludge furnace was incinerating only organics, it's phosphorus content was 1.5 mg/l, all particulate. When the lead lime centrifuge is used for dewatering

purposes only, the lime centrate is directed to the primary where the lime solids that are in the centrate settle out and are dewatered and dried or incinerated with the organic sludges. During this period the phosphorus content of the sludge furnace scrubber water increased to 3.5 mg/l, as a result of the lime centrate being directed to the primary. Since the lime centrifuge is operated with the highest possible solids capture when it is used as a dewatering device, the majority of the inert materials are conveyed to the recalcination furnace with the lime cake. This resulted in a much higher phosphorus scrubber water content of 12 mg/l. Only total ortho phosphorus was analyzed during this period. During the period of normal operations when the lead lime centrifuge is used for classification the sludge furnace, drying the concentrated waste lime inerts increased to 18 mg/l, mostly particulate. Since the majority of the inert materials had been classified out of the lime cake, the recalcination furnace scrubber water content decreased to 2.4 mg/l, mostly particulate.

Lime Mud Recalcining. Since April 1968 the District has successfully recalcined lime mud from the lime chemical treatment process. Over this period makeup lime has accounted for only 28 percent of the calcium oxide used. Average monthly CaO values in the recalcined lime have ranged between 51.0% and 74.7% with the average over the entire period being 66.0%. Table 15 shows the operating data for the lime mud recalciner. A reduction of approximately 40% in fuel requirements was achieved when centrifugal classification was used.

In an effort to measure the usable calcium losses in the recalcination furnace, the assumption was made that all the calcium lost as fly ash would be captured in the wet scrubber. The increase in calcium and phosphate in the scrubber water as a result of fly ash consisting of usable calcium and inert calcium hydroxyapatite was measured by acidifying the scrubber influent and effluent samples to pH 2.0, filtering with a .45 u filter, neutralizing the filtrate to pH 7 and analyzing for calcium hardness and phosphate. The phosphate was measured to determine the amount of calcium combined with the hydroxyapatite. The difference in terms of calcium, of the calcium hardness and the calcium combined with the hydroxyapatite provided the amount of usable calcium loss from the furnace. By measuring the amount of usable calcium in the centrifuge cake entering the furnace, described earlier in the classification and dewatering portion of this section, the percent calcium losses from the furnace can be determined. The amount of usable calcium loss from the furnace in the scrubber water was evaluated over an eight hour composite period. Using the analysis procedures and assumptions described above, 3.7% by weight of the usable calcium entering the furnace during the eight hour period was lost from the furnace.

TABLE 15

LIME RECALCINER OPERATING DATA

	<u>No Centrifugal Classification</u>	<u>Centrifugal Classification</u>
Period	Oct. 69 - May 71	July 70 - Nov. 70
Chemical Clarifier Flow, MGD	2.95	3.63
Feed, lbs/hour ⁽¹⁾	435	756
Feed, % Solids	33.3	42
Electricity KWHR/day ⁽²⁾		508
Fuel Requirements ,BTU/lb ⁽³⁾	5500	3270

(1) Pounds of dry solids per hour

(2) Includes energy and demand charges for furnace support motors and recalcined lime conveying system. February 70 to December 1970

(3) Natural gas at 18 psia and 860 BTU/ft³

Optimum Furnace Conditions. Eades and Sandberg in their discussion of lime reaction parameters point out that, "Although the art of lime burning has been practiced since ancient times, it was not until the 18th century that a scientific explanation of calcination was advanced. As industrial and chemical technology developed, lime became an increasingly important component in numerous reactions and processes. For these applications, the lime was judged primarily on its chemical purity, and minimal amounts of silica, alumina, iron, and other impurities were desired. In general, reaction rates were not thought important."

"Reaction rates of commercial limes became a matter of considerable interest with the introduction of the basic oxygen converter steel furnace. Operation costs for such converters are quite high, and steel producers quickly began investigation of methods to lower the time per heat of steel. Metallurgists became interested in the relationship between lime reactivity and slagging time. As a result of these and other studies, it quickly became apparent that in steel making and many other industrial applications reactivity of a lime was a more significant criterion for judging quality and suitability of a lime than chemical purity."

The authors further state, "that pore space and calcium oxide (CaO) crystallite size are the prime factors controlling reactivity of any given lime. These in turn can be linked to calcining conditions with low temperature burning producing a highly porous, highly reactive lime and high temperature burning producing a shrunken, dense lime with low porosity and low reactivity".

According to the AWWA Standard for Quicklime and Hydrated Lime (AWWA B202-65), high-reactive, soft-burned lime will show a temperature rise of 40°C in 3 minutes or less and the reaction will be complete within 10 minutes when tested by the method given in the standard. A medium-reactive, medium-burned lime will show a temperature rise of 40°C in 3 to 6 minutes and the reaction will be complete in 10-20 minutes. For low reactive, hard-burned lime more than 6 minutes will be required for a 40°C temperature rise and the complete reaction time will take longer than 20 minutes.

Since it is possible to produce quicklime with the same calcium oxide content, but with very different slaking properties, tests were performed on the lime recalcining furnace. The purposes of the tests were to determine the optimum recalcining temperature, feed rate and rabble rate in terms of lime reactivity and furnace fuel requirements.

The Rapid or EDTA Method for calcium oxide as described in Section XI and the AWWA Slaking Rate Test at 400 rpm were used to determine the reactive properties of the recalcined lime. At least an hour was allowed after changes in process variables to permit the furnace to reach equilibrium. A sample was then composited at 15-minute intervals from the No. 6 hearth for an additional hour. Composite feed samples were also collected at the same intervals and recalcined in the laboratory muffle furnace at 1800°F for 1 hour to insure that the potential input calcium oxide remained constant.

To establish a qualitative base line, calcium oxide and slaking rate tests were performed on virgin makeup lime. The data presented in Figure 48 shows that the makeup 16 x 50 mesh granular quicklime used at South Lake Tahoe is highly reactive. A 40°C temperature rise is reached in less than 30 seconds and the reaction completed in seven minutes. Identical results were achieved in three separate tests of the sample.

The effect of temperature on the recalcined lime activity at a constant feed and rabble rate was first investigated. Table 16 shows that recalcining temperatures between 1600°F and 1900°F had a major effect on recalcined lime activity, although all three temperatures produced lime which, under the AWWA standard, was considered to be highly-reactive. Within this temperature range there was no indication that the lime was being over-burned, since the total reaction time was well within the time requirements. Recalcining lime at 1900°F as opposed to 1800°F produced a 5% increase in available calcium oxide, but very little improvement in an already acceptable slaking rate.

At 1600°F the flour-like recalcined lime showed pronounced tendencies to agglomerate into soft, easily crushed particles of 1/4 inch to 3/4 inch diameter. Many of the particles contained centers of unburned organic sludge. Additional evidence of unburned organic sludge was observed in the Dewar Flask after the slaking test.

For the second phase, the rabble rate was varied between 1.5 and 2.0 rpm while the temperature and feed rate were held constant. At both 1900°F and 1600°F the variation of rabble rate showed very little effect on the recalcined lime activity. Once again temperature proved to be the major variable. Table 17 shows the results of varying the rabble rate.

Finally, to determine the effect of feed rate, the temperature and rabble rate were held constant at 1900°F and 1.5 rpm, respectively. The feed rate to the centrifugal was run at 870, 820, and 450 pounds of dry solids per hour. Table 18 shows that the total slaking time doubled when the feed rate was reduced from 870 lbs/hr to 450 lbs/hr with no significant change

FIGURE 48
VIRGIN LIME SLAKING RATE TEST
AWWA STANDARD B202-65

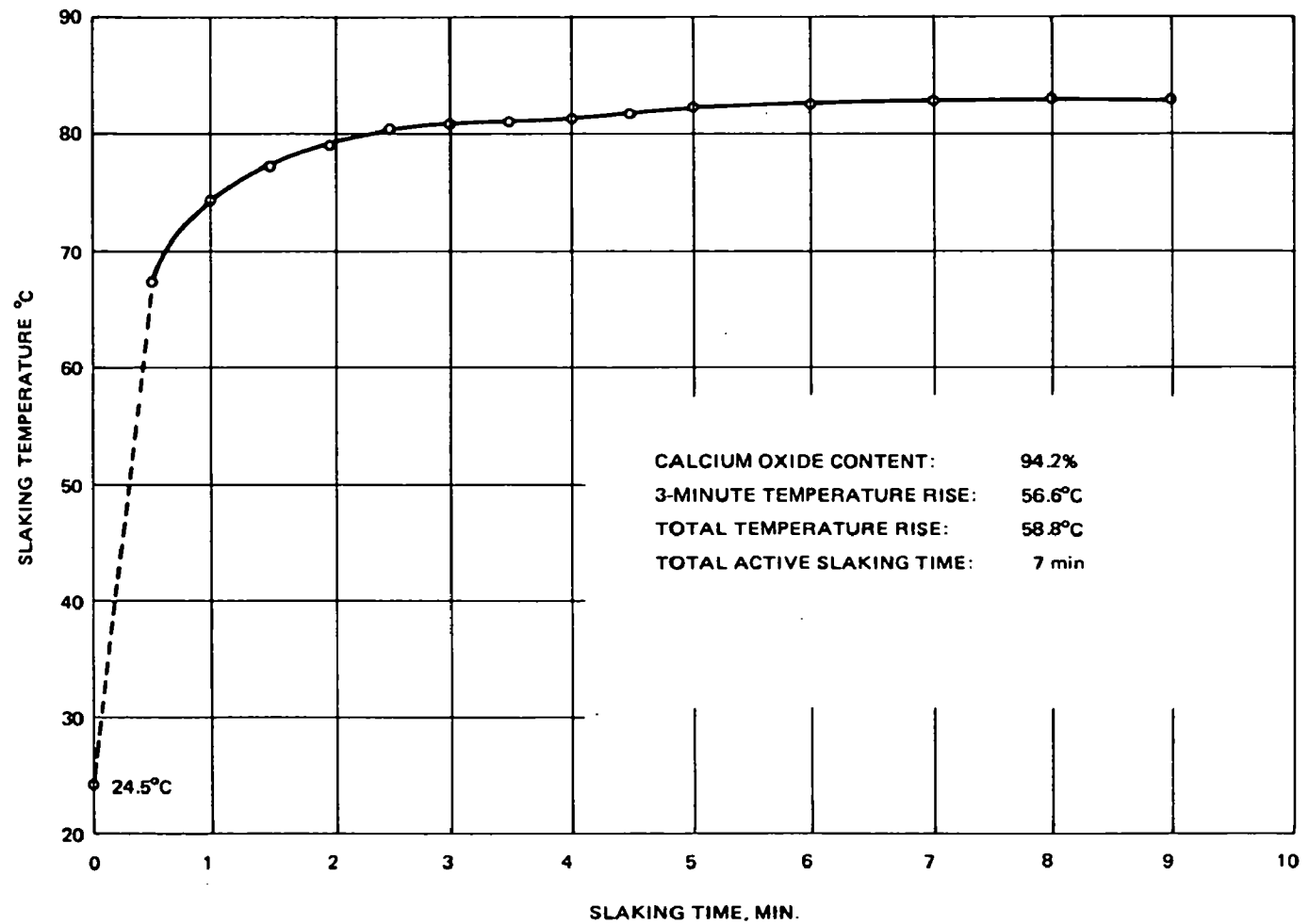


TABLE 16

Effect of Temperature On
Recalcined Lime Activity
At a Constant Feed and
Rabble Rate

	<u>Recalcining Temperature °F</u>		
No. 3 Hearth	1640	1630	1450
No. 4 Hearth	1900	1710	1620
No. 5 Hearth	<u>1900</u>	<u>1780</u>	<u>1600</u>
Percent CaO	86	81	76
Slaking Rate:			
60 Sec. Temp. Rise, °C	50	48	41
Total Temp. Rise, °C	51.5	48	41
Total Reactive Time, min.	2	1	1
Feed Rate, lbs/hr ⁽¹⁾	870	800	870
Rabble Rate, rpm	1.5	1.5	1.5

(1) Dry solids to centrifuge

TABLE 17

Effect of Rabble Rate
On Recalcined Lime
Activity At a Constant
Feed Rate

	<u>Rabble Rate, rpm</u>	
	<u>2.0 rpm</u>	<u>1.5 rpm</u>
<u>1900°F Avg Temp (1)</u>		
Percent CaO	86	86
Slaking Rate		
60 Sec. Temp. Rise, °C	51.5	50
Total Temp Rise, °C	52	51.5
Total Reaction Time, min.	1.5	2
Feed Rate (2), lbs/hr	940	870
<u>1600°F Avg Temp (1)</u>		
Percent CaO	70	76
Slaking Rate		
60 Sec. Temp. Rise, °C	39	41
Total Temp Rise, °C	39.5	41
Total Reaction Time, min.	2	1
Feed Rate (2), lbs/hr	830	870

(1) Average of No. 4 & No. 5 Hearth temperatures

(2) Dry solids to centrifuge

TABLE 18

EFFECT OF FEED RATE ON
 RECALCINED LIME ACTIVITY
 AT 1900°F ⁽¹⁾ AND 1.5 RPM
 RABBLE RATE

	<u>FEED RATE (2) lbs/hr</u>		
	<u>870</u>	<u>820</u>	<u>450</u>
Percent CaO	86	90	89
Slaking Rate			
60 Second Temp. Rise, °C	50	50.5	50.5
Total Temp. Rise, °C	51.5	53.5	53.5
Total Reaction Time, min.	2	2.5	4

(1) Average of No. 4 and No. 5 hearth temperatures

(2) Dry solids to centrifuge

in the calcium oxide content. All three feed rates produced highly-reactive recalcined lime under the AWWA standard. However, the lower feed rate produced a less reactive lime.

The optimum furnace conditions in terms of recalcined lime activity appear to be about 1900°F on the fourth and fifth hearths at 1.5 to 2.0 rpm rabble rate for 800-900 lbs/hr of dry solids to the centrifuge. The centrifuge was being used as a classifying device during this test period. On the basis of the 75% capture that was obtained, the actual furnace feed rate for the optimum conditions was 600 to 700 lbs/hr.

A slightly less reactive lime was obtained at an average recalcining temperature of 1750°F. At the same loading rates and taking into consideration the higher natural gas consumption, increasing the calcium oxide content from 81% at 1750°F to 86% at 1900°F saved approximately \$2.00 per ton of dry solids fed to the furnace.

The soft burned recalcined lime produced at South Lake Tahoe has, as previously mentioned, a flour-like texture. Individual particles are not easily seen without magnification. The large surface area to volume ratio makes the recalcined lime very easy to slake. Tables 16, 17 and 18 all indicate that the total slaking time was four minutes or less; whereas the highly reactive 16 x 50 mesh granular makeup lime required seven minutes for total slaking.

Conclusions. At the 7.5 mgd design flow about 34 tons per day of dewatered lime would have to be disposed of if lime recovery and reuse were not practiced. Through lime recalcination the total blow-down of waste solids is reduced to approximately 1.5 tons of dry solids. The cost of recalcined lime is slightly more than that of new lime, however, the reuse has and will avoid the costs of disposal and purchase or production of CO₂. Lime recovery and reused at South Lake Tahoe has demonstrated the following:

1. Lime recalcination provided not only 72 percent of the lime used for the past three years, but also a usable source of carbon dioxide.
2. Industrial gravity type thickeners are an effective device for thickening lime mud containing organic solids with very low weir overflow losses of usable calcium.
3. A concurrent flow centrifuge can be used to separate or classify phosphate rich inerts and magnesium from reusable calcium carbonate.

4. Three centrifuge bowl speeds, 2200, 1800, and 1600 rpm, were used to determine optimum bowl speed. All three speeds produced approximately the same percent cake solids and recovery or capture. Consequently the lowest bowl speed, 1600 rpm, was selected for plant operations to reduce wear and maintenance costs.

5. The lime centrifuge performed as expected with high solids captures at high pool depths and lower flows to the machine. Conversely, at low pool depths and higher flows solids capture was less.

6. At 9 tons of solids to the furnace per day, the optimum recalcining conditions were 1900°F on the number 4 and 5 hearths with a 1.5-2.0 rpm rabble rate.

7. Of the three parameters, recalcining temperature, rabble rate and feed rate, temperature had the most effect on recalcined lime activity. The CaO content in the recalcined lime was increased 15% by raising the temperatures from 1600°F to 1900°F.

SECTION XXV

CAPITAL AND OPERATING COSTS FOR CONVENTIONAL AND ADVANCED WASTE TREATMENT

Introduction. The increasing interest by the general public in the quality of the environment has stimulated several questions concerning water pollution abatement. One of the most significant questions is what the cost of cleaning up our nation's lakes and rivers will be. In some instances, the addition of secondary treatment will be sufficient to reduce the problem; however, in other areas one or more of the various advanced waste treatment processes will be required to eliminate a pollution problem. A knowledge of the costs for the various degrees of conventional and advanced waste treatment is essential in planning the nation's water pollution abatement needs.

The purpose of this section is to present in detail the actual costs of conventional and advanced waste treatment and to briefly review the benefits of the plant scale conventional-advanced waste treatment scheme used continuously since 1968 at South Lake Tahoe.

This section will show that the total cost at 7.5 mgd was \$166/mg for conventional waste treatment and \$217/mg for advanced waste treatment. These costs are based on producing the extremely high quality reclaimed water described in earlier sections with 100 percent reliability. Lesser requirements should demonstrate lower costs. Thus at South Lake Tahoe, the cost of advanced waste treatment is approximately 30 percent greater than the cost of conventional treatment.

The conventional and advanced waste treatment phases, with the exception of the ammonia stripping tower, are designed for 7.5 mgd. The ammonia stripping tower design capacity is 3.75 mgd.

Conventional treatment includes primary clarification, and both plug flow and completely mixed activated sludge secondary treatment. The activated sludge process is operated with a high organic loading and low mixed liquor suspended solids and sludge age, to prevent nitrification

and to keep the majority of the nitrogen in the ammonium ion form. The mixed liquor is chlorinated at 2 mg/l before clarification if nitrifying organisms begin to proliferate. Raw and waste activated sludges are dewatered in centrifugals and then incinerated.

Advanced waste treatment begins with phosphorus removal and clarification of the secondary effluent, using lime. The spent lime mud is thickened in a gravity thickener, dewatered by centrifuging, and then recalcined in a multiple hearth furnace for reuse. Phosphorus rich lime mud is classified in the centrifugal and wasted to the organic sludge system.

The effluent from the lime clarifier flows through an ammonia stripping tower to a two-stage recarbonation system. Scrubbed stack gases from the lime recalcining and sludge incineration furnaces are used to neutralize the high pH water. The recarbonated effluent then is pumped to mixed media filters and carbon columns. Two ballast ponds, each one million gallons in capacity, float on the system in order to provide flow equalization and supplemental filter backwash water. Spent carbon is withdrawn periodically from the carbon columns, thermally reactivated in a separate multiple hearth furnace, and then returned to the carbon columns. The carbon column effluent is dosed with 2 mg/l of chlorine and then lifted 1,500 feet and through 27 miles to Indian Creek Reservoir in Alpine County, California.

Assumption for Capital Costs. The capital costs include all equipment and construction costs. They do not include design costs. The equipment and construction costs were taken from District records of actual contracts awarded for the various treatment phases. These contracts were completed at various periods between 1960 and 1968. The EPA Sewage Treatment Plant Construction Cost Index, Base Year 1957-1959=100, was used to adjust the capital costs to 1969. It was assumed that the San Francisco Region Indexes were equivalent to South Lake Tahoe. All costs were adjusted to 1969 San Francisco Index and then to the National Average Index for 1969. The 1969 Index values used were 136.2 for San Francisco and 127.1 for the Nation. The 1969 replacement costs per million gallons were based on the national average at 7.5 mgd design capacity assuming capital amortization of all the costs at 5 percent for 25 years and no federal assistance. In fact, the capital costs to the District were much lower since federal grants from the USPHS, EDA, and EPA financed approximately 46 percent of the total construction cost for conventional and advanced waste treatment.

Capital Costs. The capital costs are shown in Table 55 and also later in the operational cost tables. Included in the capital cost figures for a specific treatment phase are items common to several processes.

TABLE 55
SOUTH TAHOE PUBLIC UTILITY DISTRICT, CALIFORNIA
CAPITAL COSTS FOR CONVENTIONAL AND ADVANCED WASTE TREATMENT PLANT
7.5 MGD DESIGN CAPACITY

Treatment Phase	Actual Contract Total Construction Cost Per Phase [2]	Estimated National Average Replacement Construction Cost for 1969	Estimated Replacement Costs Per MG for 1969
CONVENTIONAL TREATMENT			
Primary	\$ 692,000	\$ 753,000	\$ 19.50
Activated Sludge	1,247,000	1,300,000	33.60
Organic Sludge [5]	583,000	545,000	14.10
Chlorination	9,000	11,000	0.30
TOTAL, Conventional Treatment	\$2,531,000	\$2,609,000	\$ 67.50
ADVANCED TREATMENT			
Nutrient Removal			
Phosphorus Removal			
Lime Treatment	401,000	378,000	9.70
Lime Recalcining [6]	552,000	516,000	13.60
SUBTOTAL, Phosphorus Removal	\$ 953,000	\$ 894,000	\$ 23.20
Nitrogen Removal [7]			
	327,000	310,000	8.00
Recarbonation			
	162,000	152,000	4.00
SUBTOTAL, Nutrient Removal	\$1,442,000	\$1,356,000	\$ 35.20
Filtration			
	706,000	687,000	17.80
Carbon Treatment			
Carbon Adsorption [8]	656,000	632,000	16.30
Carbon Regeneration	183,000	198,000	5.20
SUBTOTAL, Carbon Treatment	\$ 849,000	\$ 831,000	\$ 21.50
TOTAL, Advanced Treatment	\$2,996,000	\$2,874,000	\$ 74.50
TOTAL WATER RECLAMATION			
Conventional Treatment	2,531,000	2,609,000	67.50
Advanced Treatment	2,996,000	2,874,000	74.50
TOTAL, WATER RECLAMATION	\$5,527,000	\$5,483,000	\$142.00

[2] Construction costs are taken from District records of actual contracts awarded for various phases. Contracts for construction were completed at various periods between 1960 and 1968. These costs have not been adjusted to a common year.

[5] Includes sludge handling, dewatering, incineration, and ash disposal.

[6] Includes lime mud handling, dewatering, and recalcining.

[7] Ammonia stripping.

[8] Includes initial carbon costs to fill all carbon columns.

These items include buildings, electrical systems, piping, and controls not specifically identifiable in the construction contracts. Both area and volume relationships are used to proportion this capital cost to specific treatment phases.

Total capital costs, based on 1969 national average replacement costs, were 67.50 \$/mg for conventional treatment and 74.50 \$/mg for advanced treatment. The 20 acre site where the present conventional and advanced waste treatment processes are located was acquired in conjunction with the construction of the original 2.5 mgd primary and secondary plant in 1960. Since the records did not show the site acquisition as a separate item, the cost of the 20 acres was included only in the capital cost of conventional treatment.

Assumptions for Operating Costs. The operating costs included in this section are based on the plant design capacity of 7.5 mgd from February 1969 to December 1970. During this period, the actual average monthly influent flows varied between 1.79 mgd and 3.15 mgd. As a result of recycling water from scrubber flows, backwashing filters, and backflowing carbon columns, the filtration and carbon adsorption flows during the same period varied between 3.13 mgd and 5.22 mgd.

To compare operating costs on the common basis of the plant design capacity, it was assumed the total cost for fuel, chemicals, make-up lime, and make-up carbon would increase in proportion to the flow. However, the same assumption could not be made for electricity, operating and maintenance labor costs, equipment repair, and instrument maintenance. Costs per day for electricity were adjusted upward to reflect the equipment characteristics at 7.5 mgd. The cost per day for operating and maintenance labor, repair materials, and instrument maintenance were assumed to be the same as at present and design flows.

The assumption that the plant is staffed as though it were operating at design capacity was made for the following reasons. During the grant, three individuals were required per shift to operate the plant. One operator spent approximately five hours in the laboratory analyzing samples for the District's FWQA research grant, and three hours performing plant related duties. The second operator spent an hour per shift measuring flows and keeping track of expendables for the grant, and seven hours in plant operation. The third individual controlled the operations within the incineration building, which did not vary with flow. This extensive data collection would not be taking place at 7.5 mgd nor would it be the usual practice at other advanced waste treatment plants. Maintenance labor and repair materials would be affected by equipment age but not necess-

arily by flow since most of the equipment is running today. Because it was the basic purpose of this section to show costs at 7.5 mgd in 1969 and 1970, age was not considered to be a factor.

It was assumed, however, that emphasis placed on each treatment phase by the first two operators would be different during carbon regeneration periods when this additional phase must be covered.

No District operating and capital costs associated with sewers, pump stations, janitorial work in administrative areas, minor plant modifications, effluent export, FWQA research grant, and general Utility District administration were included. It is felt that these costs were not typical operating costs for actual waste treatment.

Operating Cost Data Collection and Analysis. Each month the computer provided machine listings of the individual operating costs per treatment phase for current and design plant flows. The computer also determined and printed the average to date costs for the current and design flows.

The monthly labor rates for the operations and maintenance included the actual monies paid by the District to or in behalf of the individual for straighttime, overtime, standbytime, holidays, vacation, sick leave, premium pay, Social Security, retirement fund, medical insurance, unemployment tax, and Workmen's Compensation. The labor rates represented all monies paid divided by the actual hours worked.

Two methods were used to allocate labor costs to the various treatment phases. For maintenance labor, the actual hours spent within a specific treatment phase were fed into the computer. Overhead time, such as supervision, lunch, coffee breaks and unreported time, was prorated among the various treatment phases on the basis of actual hours reported. The maintenance group included six individuals.

Operating labor was divided among the various treatment phases by fixed percentages. Three shifts were used, seven days per week. The basic shift included three operators. A total of 15 operators were needed to cover the three shifts, chief operator, and vacation and sick leave make-up. The fixed percentages used to allocate the operational labor hours are shown in Table 56. The fixed percentages were based on personal interviews with each of the operators, and on observations by the authors and the District administrative staff. Again, these percentages were stored in the computer as variable constants.

TABLE 56

PERCENT OPERATIONAL LABOR DIVISION
PER TREATMENT PHASE

<u>Treatment Phase</u>	<u>Without Carbon Regeneration</u>	<u>During Carbon Regeneration</u>
PLANT EXCLUDING INCINERATION BUILDING - 2 Operators/Shift		
Primary Treatment	20.80	13.53
Secondary Treatment	25.00	17.70
Lime Clarification	4.17	.83
Ammonia Stripping	2.10	.42
Recarbonation	2.08	.83
Filtration	6.25	4.17
Carbon Adsorption	6.25	4.17
Carbon Regeneration		25.00
INCINERATION BUILDING - 1 Operator/Shift		
Organic Sludge Dewatering	6.25	6.25
Organic Sludge Incineration	4.17	4.17
Lime Mud Dewatering	4.17	4.17
Lime Mud Recalcining	12.50	12.50
Lime Clarification (Slakers)	4.17	4.17
Recarbonation (CO ₂ Compressors)	2.09	2.09
TOTAL	100.00	100.00

Operator responsibilities include process monitoring and adjustments, chemical mixing, routine equipment servicing, and inside cleanup.

Electrical costs were divided among the treatment phases on the basis of ampere-hours per month. The actual running time for each electrical motor was one of the inputs for each month. The amperes for each motor were stored in the computer as variable constants.

Cubic feet of natural gas used by the sludge incinerator, lime recalciner and carbon furnace and the amounts of chemicals used each month were each measured separately.

Instruments were maintained by an outside contract. The costs per treatment phase were determined by the number of instruments repaired or calibrated per treatment phase. The contract included labor, parts, and instrument replacement.

Repair material costs included replacement costs, material, repair equipment purchase, and rental costs. This category also reflected outside labor cost, as well as material costs for work done outside the plant, such as rewinding electric motors.

Unit Operating Costs. Costs of all commodities at Lake Tahoe are high in comparison to the average of commodity costs in the USA. This is due in large measure to the location in a mountainous area and to the Tahoe area's tourist-based economy. In order to permit comparison with costs which might be anticipated in other areas, Table 57 has been prepared to show the present cost of various commodities at Lake Tahoe. These unit costs were stored in the computer as variable constants.

Primary Treatment. Raw sewage entering the plant passes through a barminutor, a parshall flume and then into the primary clarifier. A rectangular 2.7 mgd clarifier equipped with water spray scum collection and a 4.8 mgd, 100 foot diameter circular clarifier with mechanical scum collection can be used. The underflow is degrittied with a cyclone degritter.

Table 58 shows the operating and capital costs for primary treatment. The barminutor, primary clarifier, degritter, and sludge and scum withdrawal pumps were considered to be part of primary treatment.

Secondary Treatment. The primary effluent flows by gravity to the activated sludge secondary treatment system. This system consists of three plug flow aeration basins (0.9 mgd each), two completely mixed aeration basins (2.4 mgd each) all in parallel, and two circular secondary clarifiers, 2.0 and 5.5 mgd, respectively. Activated sludge is wasted to

TABLE 57
UNIT COSTS⁽¹⁾ 1969 AND 1970

Labor⁽²⁾

Operations	\$ 6.11 /hour
Maintenance	5.05 /hour
Electricity ⁽³⁾	12.10/1,000 kwh
Fuel ⁽⁴⁾	0.0543/therm
Make-up Lime (Quicklime) ⁽⁵⁾	28.83/ton CaO

Chemicals

Chlorine	114.00/ton
Polymer - Sludge Dewatering	2.53/lb
Polymer - Lime Coagulation	1.92/lb
Polymer - Filtration	1.92/lb
Alum - Filtration (Liquid) ⁽⁶⁾	0.014/lb
Activated Carbon Make-up ⁽⁷⁾	0.305/lb

- (1) All appropriate unit costs are f.o.b. South Lake Tahoe and include a 5% California sales tax.
- (2) Labor costs include all direct and indirect monies paid upon the employees behalf. The rates are averages for 1969 and 1970.
- (3) Includes energy and demand charges.
- (4) Natural gas at about 860 BTU/cu ft at 6,200 feet elevation and billed on the basis of interruptable service.
- (5) Average calcium oxide content 93.6%.
- (6) Liquid alum weight 11.08 lbs/gal. Dry alum equivalent 49%.
- (7) Activated carbon at 30 lbs/cu ft.

Overall Costs - Organic Sludge Dewatering, Incineration, and Disposal. The operating and capital costs shown in Table 62 represent the total cost for organic sludge handling and disposal excluding dumping charges.

Lime Coagulation. Chemical coagulation (with lime) of the secondary effluent to pH 11 is accomplished with a rapid-mix flocculation basin, followed by a 100 foot diameter conventional clarifier. To reach pH 11 for the South Lake Tahoe wastewater, a lime dose of 300 mg/l of calcium oxide is required. A polymer, at a 0.1 to 0.3 mg/l dose, is added just as the water leaves the flocculation chamber to improve clarification.

The lime coagulation system typically removes 95 percent of the phosphorus it receives. In the clarifier effluent, phosphorus concentrations range between 0.2 and 0.7 mg/l $\text{PO}_4\text{-P}$ and turbidity levels between 1 and 10 SJU.

Table 63 shows the specific operating and capital costs for this treatment phase. The lime storage bins, slakers, floc basin, clarifier, polymer feed system, and sludge draw-off pumps were considered to be part of the lime coagulation system. As shown in Table 56, operating labor from both the general plant and the incineration building were charged to lime clarification. Maintenance labor and repair material costs for 1969 and 1970 represented primarily costs associated with the slakers, lime buildup on the flash mixer, sludge removal from the lime mixing basin, and cleaning of the lime slurry line from the slakers to the flash mixing basin.

Lime Mud Dewatering. Lime mud is pumped from the chemical clarifier and reaction basin to a gravity thickener for solids concentrations. In turn, the thickened mud is pumped by either of two variable speed pump units to a 24-inch x 60-inch solid bowl centrifugal for dewatering. Phosphorus is wasted from the system in the form of calcium hydroxyapatite by operating the lime centrifugal so that 10-30 percent of the solids entering the machine come out in the centrate. A second centrifugal clarifies the centrate and the phosphorus-rich cake is conveyed to the organic sludge furnace.

Operating and capital costs for the dewatering phase of lime recalcination are shown in Table 64. The items of equipment included in this phase were the thickener, lime sludge pumps to the centrifugal, the centrifugal itself and lime cake conveyor to the furnace. In 1969 and 1970, most of the maintenance labor and repair costs were related to lime mud line cleaning and centrifugal repair. The dewatering costs also included the second centrifugal when it was being used to dewater lime centrate for

TABLE 63

OPERATING AND CAPITAL COSTS
LIME COAGULATION AT 7.5 MGD

<u>OPERATING COSTS</u>	<u>\$/DAY</u>
Electricity	5.22
Make-up Lime	168.19
Polymer	22.18
Operating Labor	30.67
Maintenance Labor	9.59
Repair Materials	3.14
Instrument Maintenance	<u>3.04</u>
Total Operating Cost	<u>242.03</u>
 <u>TOTAL COSTS PER MG</u>	 <u>\$/ MG</u>
Operating	32.27
Capital	<u>9.70</u>
Total	<u>41.97</u>

TABLE 64

OPERATING COSTS
LIME MUD DEWATERING AT 7.5 MGD

<u>LIME MUD DEWATERING</u>	<u>\$/DAY</u>
Electricity	8.92
Operating Labor	21.47
Maintenance Labor	16.04
Repair Materials	8.81
Instrument Maintenance	<u>0.00</u>
Total Operating Cost	<u>55.24</u>
 <u>TOTAL OPERATING COST</u>	
Per MG Plant Influent	\$ 7.37/MG
Per ton CaO Recalcined	5.49/ton CaO

the lead centrifugal.

The costs for drying and disposing of the dewatered high phosphate lime mud were included in the costs for organic sludge incineration and ash disposal.

Lime Mud Recalcining. Dewatered lime mud is conveyed from the centrifugal to a 14.3 foot diameter, six hearth furnace for recalcining at 1800-1900°F. The maximum capacity of the furnace is about 20 tons of dry solids per day. The recalcined lime exits by gravity through a crusher into a thermal disc cooler, and then is pneumatically conveyed to the recalcined lime storage bin for eventual reuse.

For the purpose of determining operating and capital costs, all equipment and functions from the lime furnace to the lime storage bin were considered to be part of the lime recalcining system. These costs are shown in Table 65. The majority of the maintenance costs were associated with the conveying of recalcined lime from the furnace to the storage bin.

Overall Costs - Lime Dewatering and Recalcining. The operating and capital costs shown in Table 66 represent both dewatering and recalcining costs. If lime mud dewatering costs are considered common to any lime clarification process, then the costs of lime recalcining are competitive with buying new lime, particularly when disposal costs are included. At South Lake Tahoe new lime was purchased at \$28.83/ton CaO, whereas lime recalcination cost \$31.61/ton CaO.

Nitrogen Removal by Ammonia Stripping. Following lime clarification, ammonia stripping is utilized at South Lake Tahoe to remove the nutrient nitrogen from the wastewater. The stripping process includes two constant speed pumps, a cross flow cooling tower with a two-speed reversible 24-foot fan, a concrete collection basin below the tower, and a flow measurement weir on the basin exit. The tower has an average design air-to-water ratio of 250 cu ft/gal, and a nominal capacity of 3.75 mgd.

Tower removal efficiencies have varied from 30 to 90 percent; depending on air temperature and extent of calcium carbonate buildup on the fill before cleaning. During the winter when the air temperature is lower than 32°F (0°C.), the tower is bypassed to prevent ice buildup on the fill. Tower influent NH₃-N concentrations have ranged from 15 to 30 mg/l, with effluent values from 3 to 15 mg/l.

TABLE 65

OPERATING COSTS

LIME MUD RECALCINING AT 7.5 MGD

<u>LIME MUD RECALCINING</u>	<u>\$/DAY</u>
Electricity	6.55
Natural Gas	142.83
Operating Labor	64.36
Maintenance Labor	12.68
Repair Materials	8.31
Instrument Maintenance	<u>3.63</u>
Total Operating Cost	<u>238.36</u>
 <u>TOTAL OPERATING COST</u>	
Per MG Plant Influent	\$ 31.78 MG
Per Ton CaO Recalcined	23.70/ton CaO

TABLE 66

OPERATING AND CAPITAL COST
LIME MUD DEWATERING AND RECALCINING
AT 7.5 MGD

<u>LIME MUD DEWATERING</u>	<u>\$/MG</u>	<u>\$/Ton CaO</u>
Operating Costs	7.37	5.49
Capital Costs	<u>2.90</u>	<u>2.16</u>
Total	<u>10.27</u>	<u>7.65</u>
 <u>LIME MUD RECALCINING</u>		
Operating Cost	31.78	23.70
Capital Cost	<u>10.60</u>	<u>7.91</u>
Total	<u>42.38</u>	<u>31.61</u>
 <u>TOTAL OPERATING AND CAPITAL COSTS</u>	 <u>52.65</u>	 <u>39.26</u>

SECTION XXVII

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LIME RECOVERY AND REUSE.

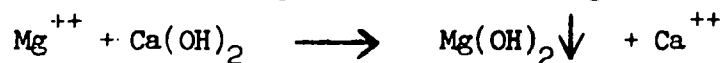
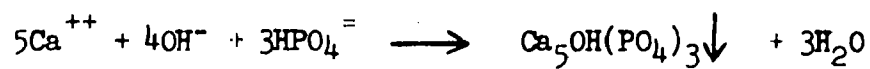
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NITROGEN REMOVAL.

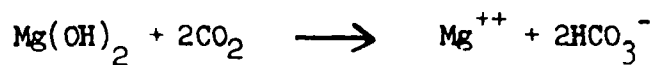
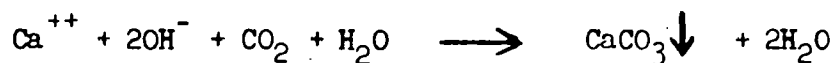
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LIME CLARIFICATION, RECOVERY AND REUSE

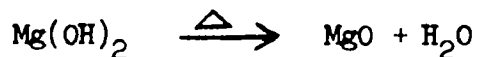
Precipitation



Recarbonation



Recalcination

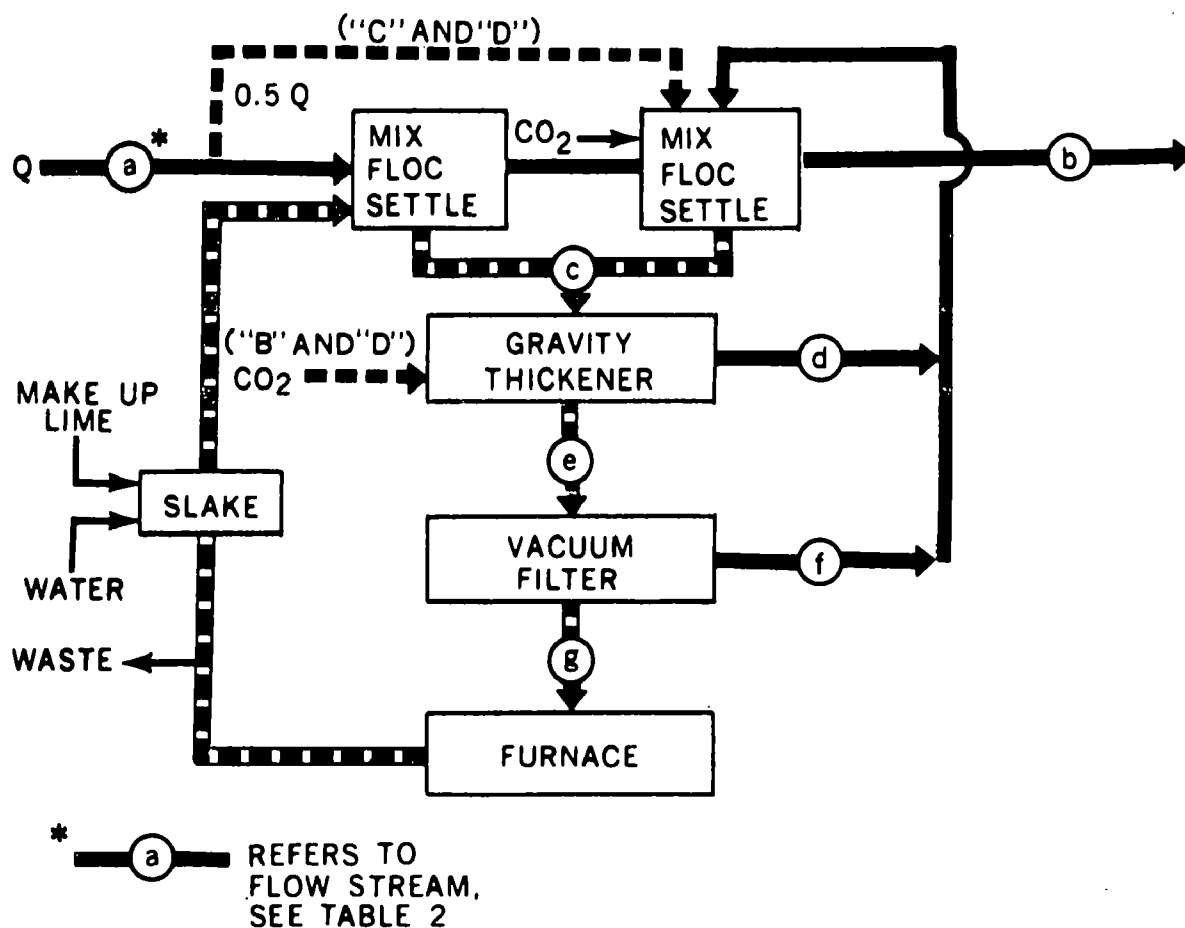


Hydration



TABLE 1

FIGURE 1
EXPERIMENTAL SYSTEMS



SYSTEM "A"
TOTAL TREATMENT

SYSTEM "B"
TOTAL TREATMENT WITH SLUDGE CARBONATION

SYSTEM "C"
SPLIT TREATMENT (0.5Q to 2nd STAGE)

SYSTEM "D"
SPLIT TREATMENT WITH SLUDGE CARBONATION

NOTE: LIME DOSE FOR SYSTEMS "C" AND "D" IS $\frac{1}{2}$ OF
THE LIME DOSE USED FOR SYSTEMS "A" AND "B"

FIGURE 2

LIME REQUIREMENT FOR $\text{pH} \geq 11.0$ AS A FUNCTION
OF THE WASTEWATER ALKALINITY

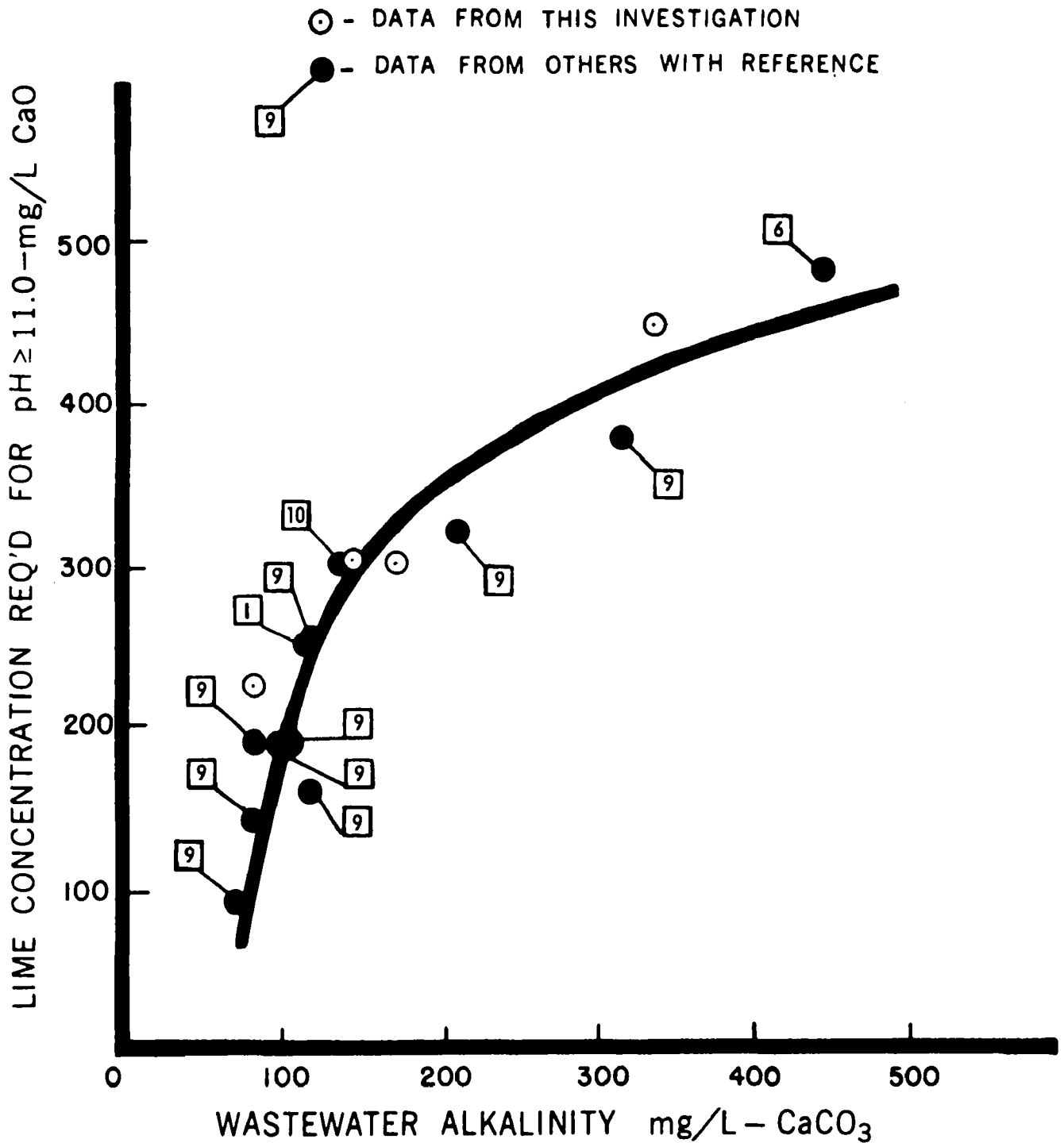


FIGURE 4

SLUDGE THICKENING CHARACTERISTICS
VS.
MAGNESIUM HYDROXIDE / CALCIUM CARBONATE RATIO

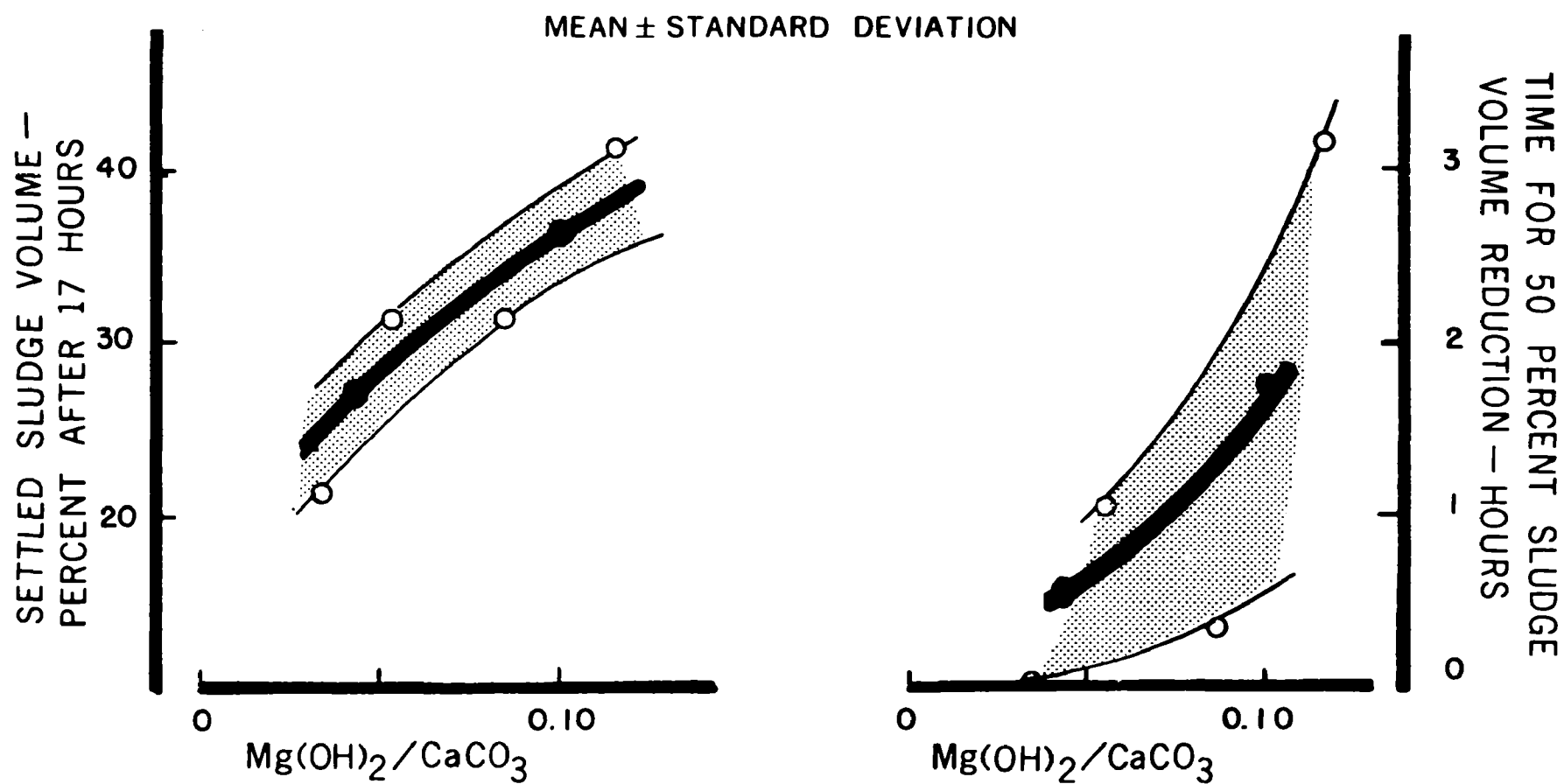


FIGURE 5

LIME DOSE vs. FILTER YIELD

MEAN \pm STANDARD DEVIATION

IMMERSION FRACTION = 0.25

VACUUM = 15" Hg

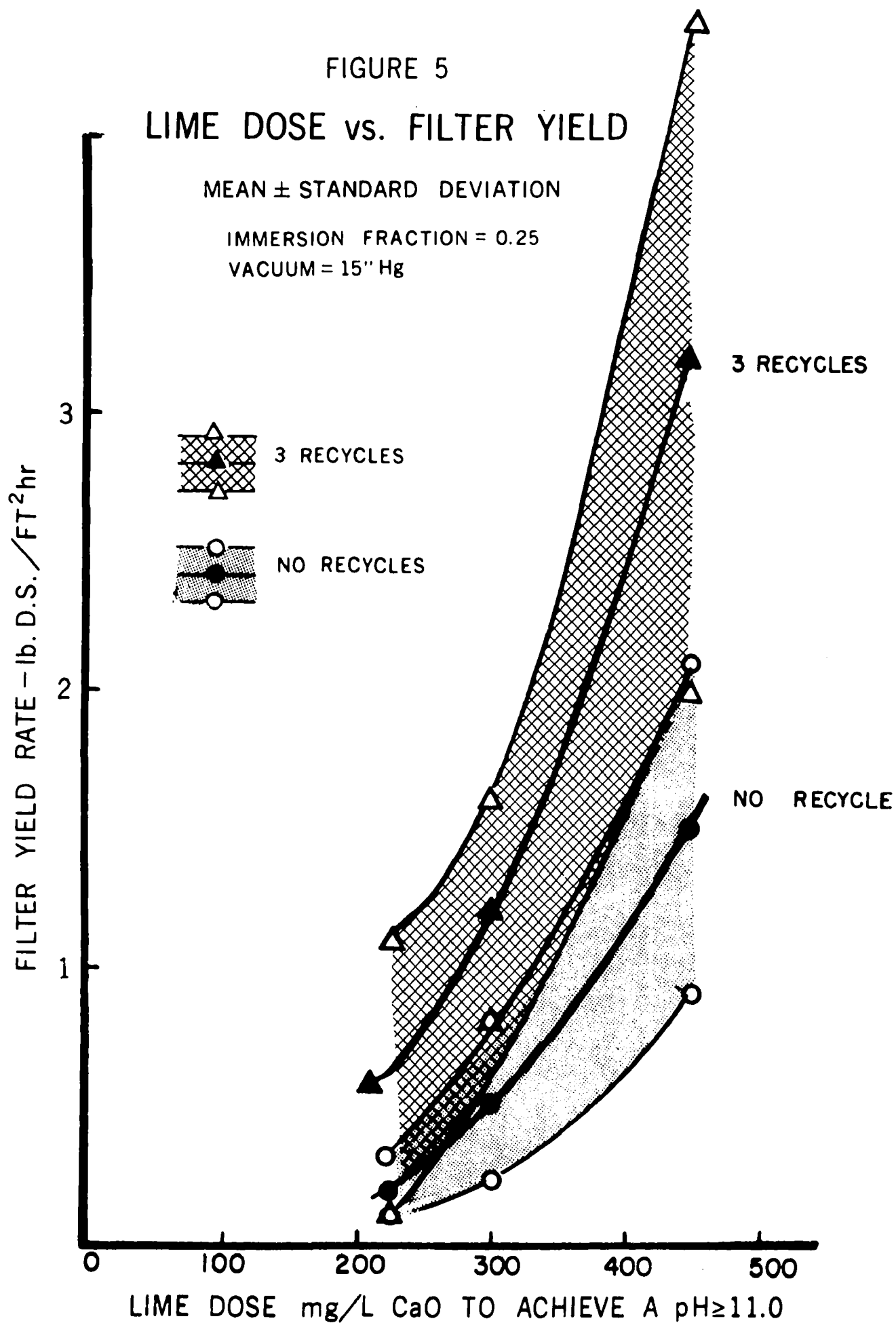


FIGURE 6A
SYSTEM "A"
TOTAL TREATMENT

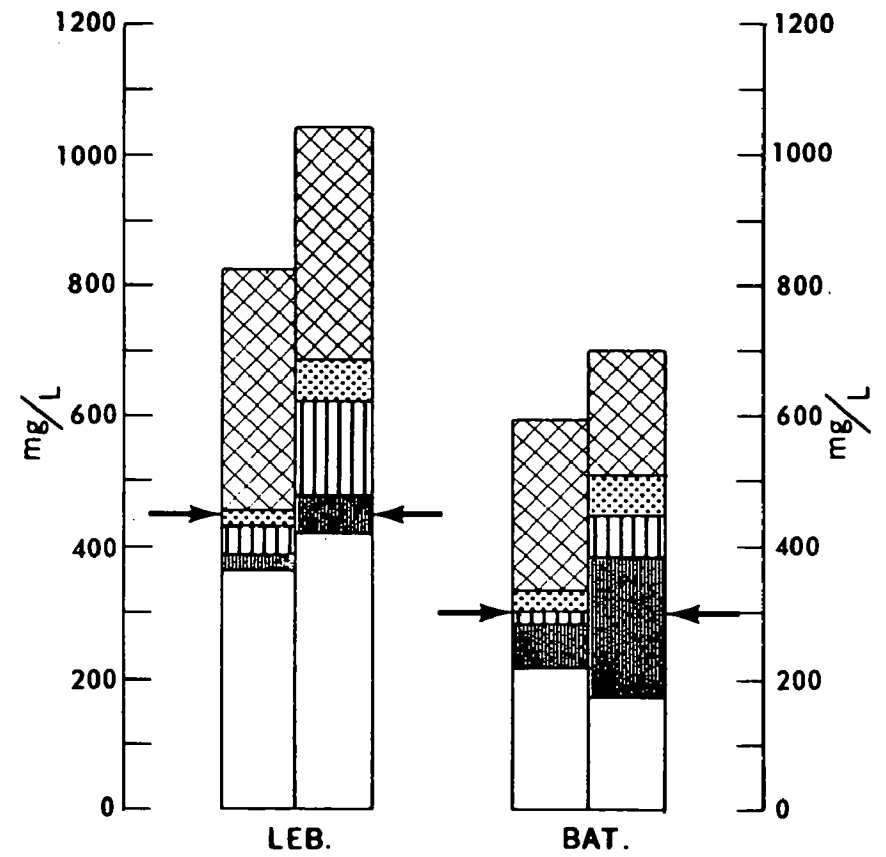
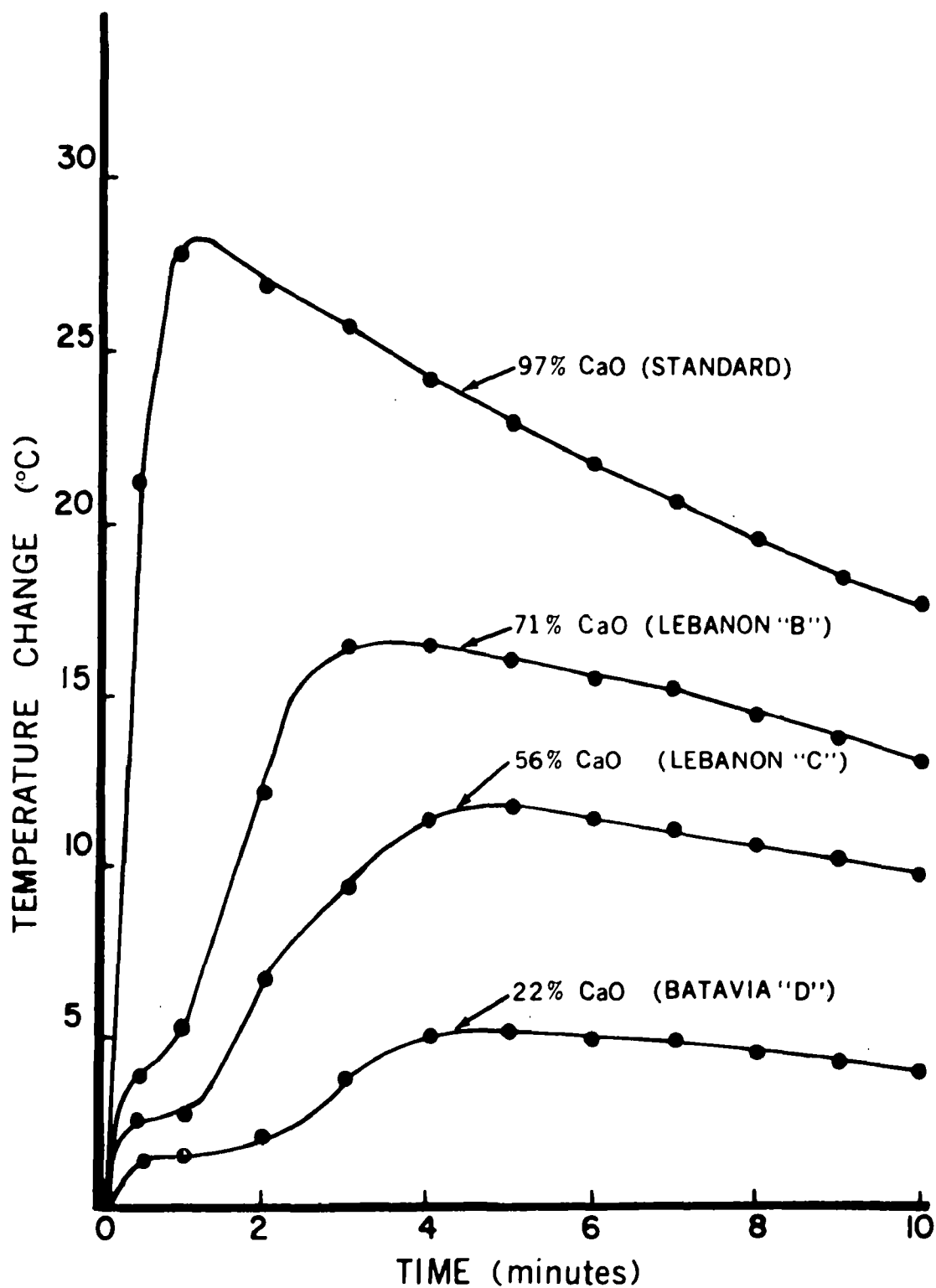


FIGURE 7
SLAKING CHARACTERISTICS OF RECOVERED
LIME AFTER THREE RECYCLES



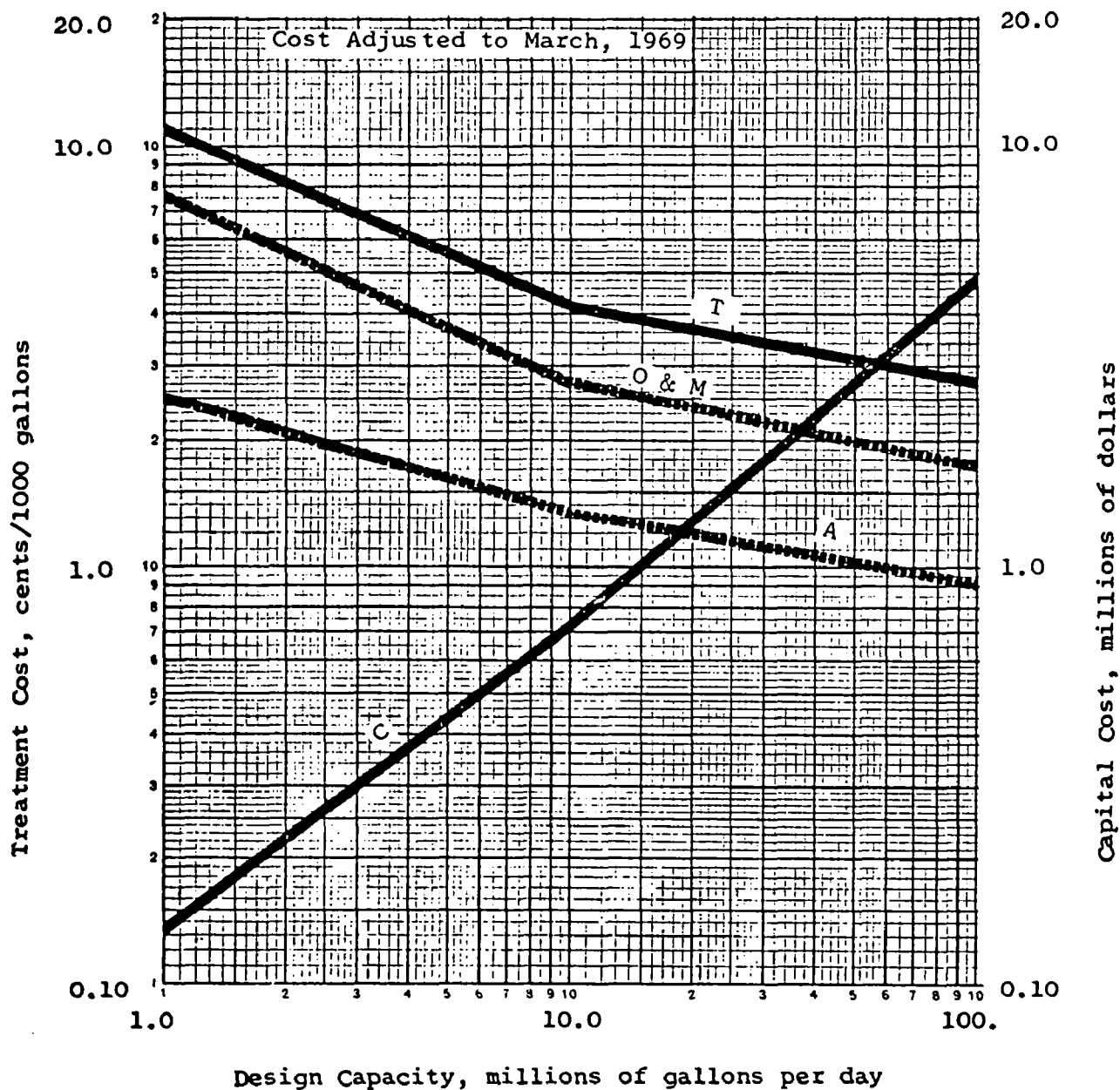
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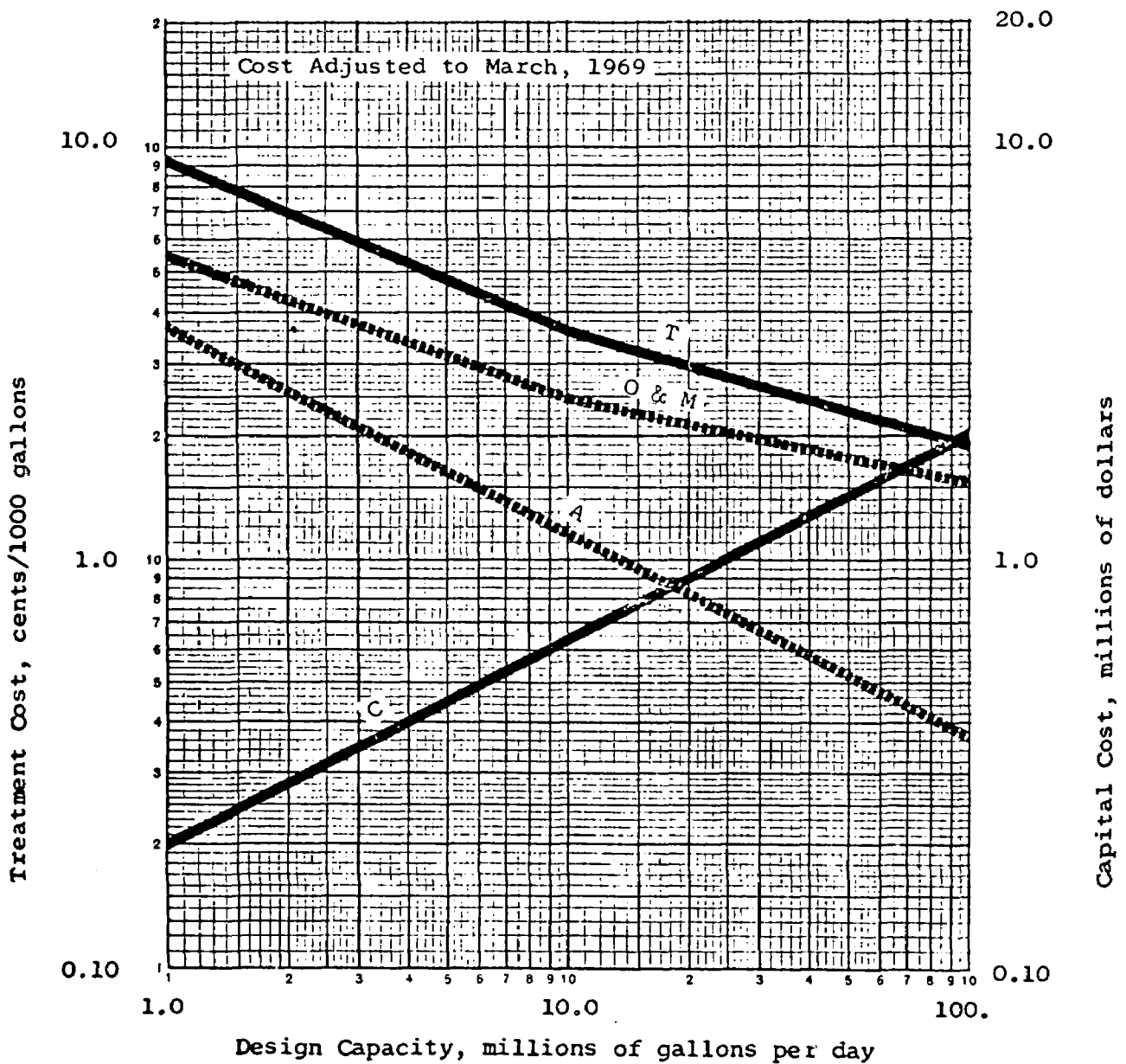
TWO CLARIFIER LIME CLARIFICATION PROCESS
WITHOUT CHEMICALS

Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity



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

LIME RECALCINATION PLUS MAKE UP LIME
FOR USE WITH LIME CLARIFICATION
Capital Cost, Operating & Maintenance Cost, Debt Service
vs.
Design Capacity



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

Figure 5

FILTRATION

CHAPTER 9

DEEP BED FILTRATION

9.1 General

With the exception of gravity sedimentation, deep-bed filtration is the most widely used unit process for liquid-solids separation. Until recently its use was generally confined to the treatment of municipal and industrial water supplies. The primary reason for its recent adoption in wastewater treatment has been the need to upgrade effluents from conventional treatment plants. Installations may use direct filtration of activated sludge or trickling filter effluents, without the addition of chemical agents. Also, deep-bed filters are employed in systems for phosphorus removal from secondary effluents and in physical-chemical systems for the treatment of raw wastewater. In these latter cases chemical coagulation, flocculation and sedimentation precede the filters as in water treatment plants.

In essence, the unit process of deep-bed filtration encompasses exhaustion of the bed followed by a regeneration. Water containing suspended solids is passed through a bed of granular material resulting in deposition of the suspended solids in the bed. Eventually the pressure drop across the bed becomes excessive or the ability of the bed to remove suspended solids is impaired. Thereupon filtration is stopped and the bed is cleaned prior to being placed back in service.

9.2 Filter Design

At the present time virtually all deep bed filters utilized for waste treatment are “rapid sand” type, i.e., downflow, static bed, with batch or semicontinuous operation. In this section, design information on this traditional type of filter will be reviewed. In a later section of this chapter, descriptions and design information for some new concepts in deep bed filtration will be given.

Figure 9-1 (3) is a cut-away view of a typical “rapid sand type” filter, illustrating most of its components. In essence the filter is a box containing filter media, an underdrain system, a backwash system, flow control systems and various conduits for bringing feedwater and wash water to and conveying filtrate and used wash water away from the filter. Figure 9-2 illustrates a pressure filter. It can be seen that there is little difference between the pressure and gravity flow filters except for the pressure housing. Because of size restrictions on pressure filters they are equipped with a simpler wash water collection trough system than gravity filters.

9.2.1 Design Parameters

Process design of a filter includes determination of:

- a. type and size of filter media
- b. depth of filter
- c. rate, duration and timing of water backwash, air scour and surface wash

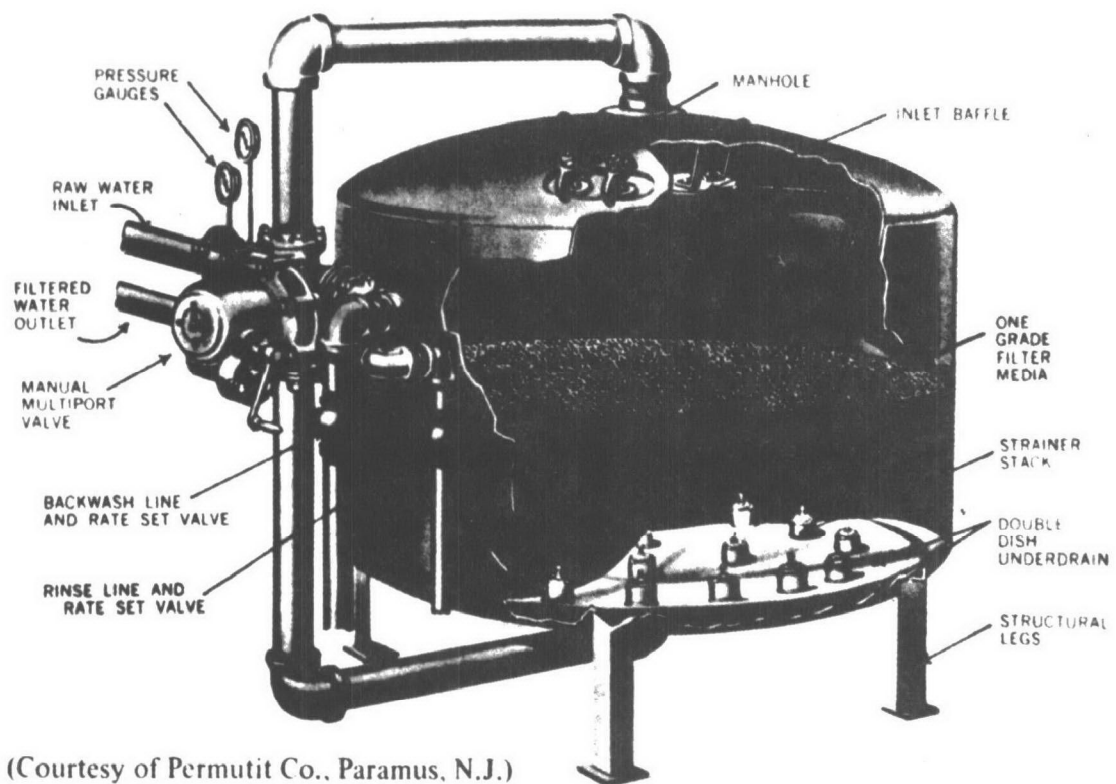
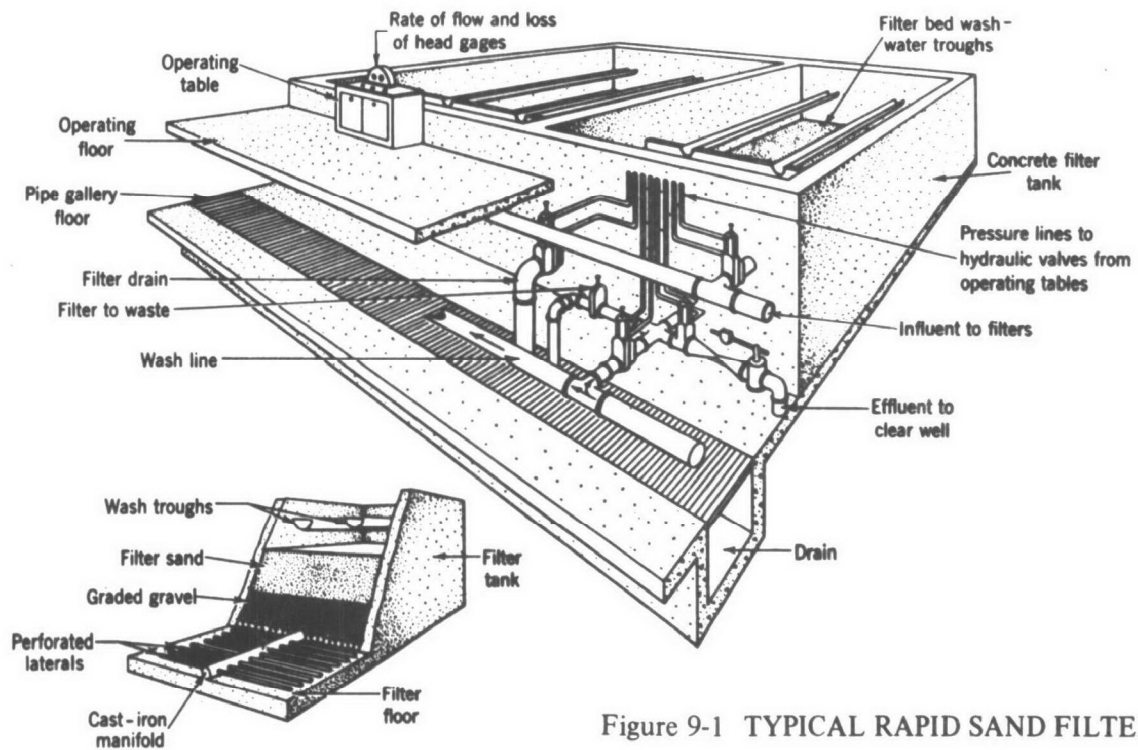


Figure 9-2 PRESSURE FILTER

- d. filtration rate
- e. type of chemical pretreatment and dose requirement
- f. expected duration of filter runs.

Values for each of these parameters must be selected to yield a design which will produce an effluent of the desired quality at a minimum cost.

The difficulty in arriving at an optimum process design is twofold. Many, if not all, of these parameters are interdependent. Also, the present level of filtration theory can only semi-quantitatively represent the interdependence of the process design parameters.

Process design information should be generated from pilot plant data. At best, present filtration theory can only reduce the degree of pilot plant study required. All too often process design of a filtration installation is performed according to "rule of thumb" experience. Use of such a procedure usually results in an overdesign (uneconomical) or an underdesign (system failure to meet quality or quantity objectives.)

9.2.2 Pilot Plant Studies

A rather large number of interdependent parameters must be considered in pilot plant studies. In addition, the range of variation of each which may have to be investigated is quite large. Thus, a field pilot investigation cannot generally be of short duration. Pilot studies up to a year in length may be required, but it should be possible to decrease this time period by utilizing experience as a guide in establishing limits of variation of the parameters.

Pilot studies can be physically described as consisting of a series of filter columns using an available supply of feedwater. Provisions must be made to measure effluent quality with time, pressure drops over various sections of the bed and over the entire bed with time and backwash rates required to achieve specified degrees of media expansion. The studies should be conducted at several flow rates, with several media types and size ranges, to various terminal headloss values and with a variety of chemical pretreatments where appropriate.

The data resulting from these studies should be utilized to eliminate from further consideration those situations which obviously cannot meet effluent criteria or cannot be justified economically. Based on these analyses, a narrowed-down set of parameters can be used in a second set of investigations. This second study may be more comprehensive than the first series, including such variables as expected diurnal variation and slight changes in chemical dose. This process may continue through several more rounds before an "optimum design" is achieved. This final design should be given a lengthy test (one month or more) at conditions as close as possible to those expected in the field. Even after this lengthy study, it is probable that only a good approximation of field conditions will be obtained.

Specific details for conducting filtration pilot studies have not been established. The following is a list of equipment and practices usually used:

- a. Multiple filter tubes of transparent material with a minimum diameter of 2 to 3 in. are utilized.

- b. The tubes are fitted for either gravity or pressure operation.
- c. A false bottom underdrain is utilized with either a porous plate or strainer backwash system.
- d. Flow control is established by the use of a positive displacement pump or an effluent throttle valve connected to a float.
- e. Pressure taps are provided above and below the media, as well as at other locations within the bed. Tap locations are generally located near the top of each type of media used.

Additional details of pilot filters are given in various references on filtration studies (1), (2).

Two areas of process design cannot be as adequately explored in pilot studies as the others listed above. These are the parameters associated with cleaning the bed and the effect of chemical treatment of the feed. Cleaning of the filter bed is difficult to simulate in pilot scale because of the small surface area of the beds utilized. The small area makes it impossible to study surface wash and air scour. Results of water backwash may not be representative because of the wall effect. Chemical treatment of the filter feed without flocculation and sedimentation may be simulated in the pilot studies. If flocculation and sedimentation are employed after chemical treatment however, the performance of model flocculators and clarifiers may not closely simulate that of full-scale field units. Thus the feed to the filters may not be representative. Both areas of difficulty can be overcome by using large pilot installations, i.e., minimum filter size of about 2' x 2' to study backwash, and minimum clarifier size of about 10' diameter to simulate full-scale sedimentation. However, going to this scale is not usually economically justified. If studies are being conducted at an existing plant, i.e., one that has full-scale clarifiers or filters, advantage should be taken of the situation.

Although pilot studies give the most reliable information on which to design filters, it should be remembered that filtration is a mixture of art and science. Excessive fine tuning of the process design should not be attempted. The chosen process design should provide a flexible system which will do the job with an adequate safety factor at a reasonable cost. Fine tuning of the process should be left to the plant operator.

9.2.3 Filter Media

Selection of the size, type and depth of the filtration media is the single most important decision in the design of a filtration system. Unfortunately, it is impossible to specify optimum media characteristics from theoretical considerations. Pilot studies should be conducted with several sizes and types of media prior to making a decision.

Sand has historically been the filtration medium most commonly used, but anthracite coal and, to a lesser extent, garnet have been employed. These substances occur in nature and are used in filters in a graded size range. Although use of a uniform size medium would have certain advantages, it is not economical to excessively restrict the size range.

A typical grain size distribution curve for a naturally occurring filter medium is given in Figure 9-3. This curve is often a straight line on log-probability paper. Historically, the two points used to characterize a medium are the 10 percent size and the 60 percent size. These

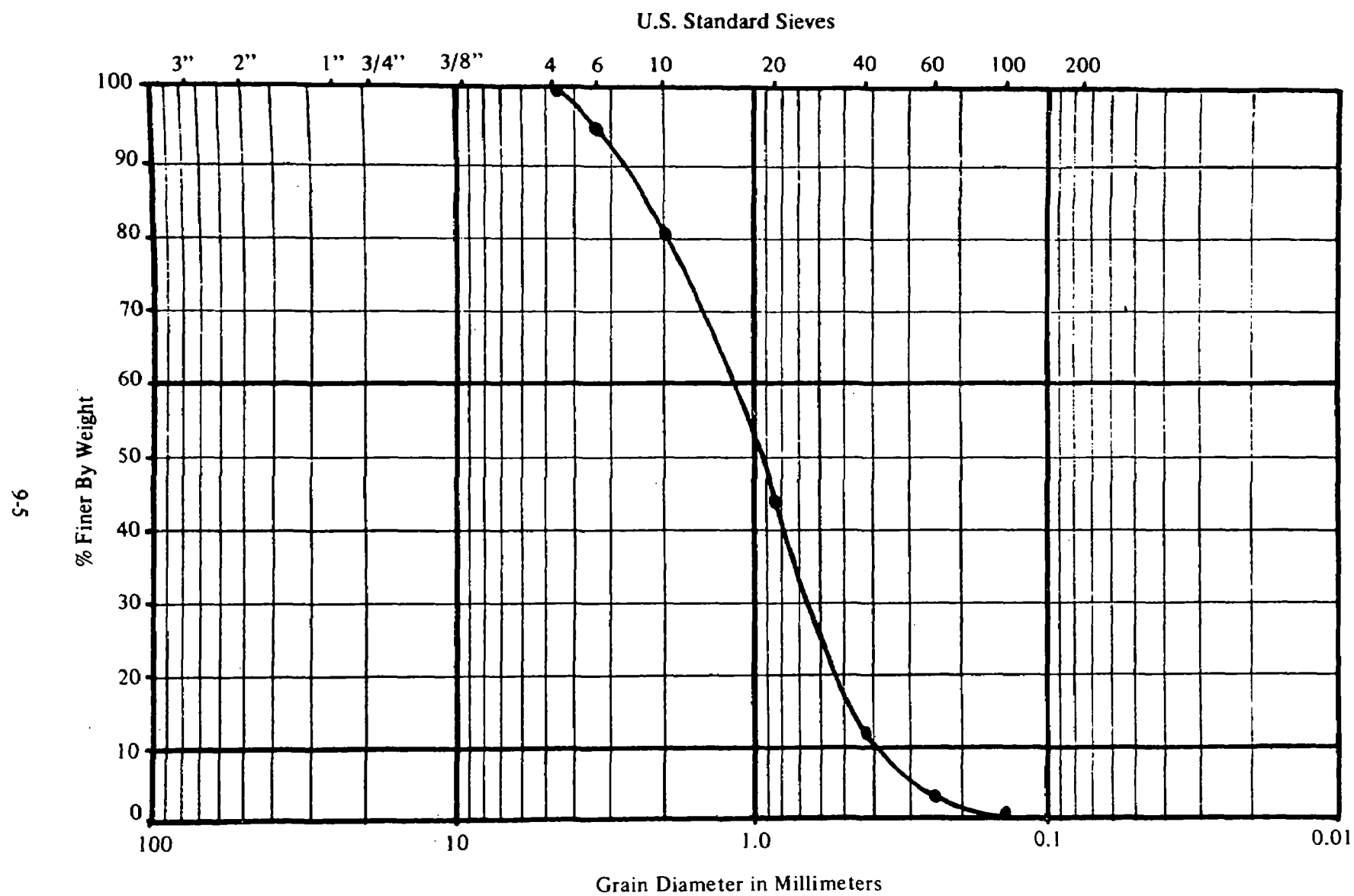


Figure 9-3 GRAIN SIZE CURVE

are defined as the particle size of a distribution such that the weight of all smaller particles constitutes the stated weight fraction of the whole. Usually the specification is given as the effective size (10 percent size) and the uniformity coefficient (ratio of the 60 percent size to the 10 percent size). In general, good correlation has been found between clean water headloss and effective size (3). Fair and Geyer (3) present a method of calculating the size fractions which must be sieved or washed out to convert from one size distribution to another.

Until recently most filter designs called for a single filtration medium, either sand or coal. Typical effective sizes range from 0.35 mm to 0.8 mm with uniformity coefficients between 1.3 and 1.7 (4). These size specification ranges can be utilized for preliminary selection of sizes to be evaluated in pilot studies.

In coal and sand media of identical size, it has been found that the solids removal with coal is somewhat inferior, but the rate of pressure drop build-up is lower (5). This finding is explained by the greater angularity of coal with consequent greater porosity.

The relationship of medium size to filter performance can be generalized. Smaller media are more effective in removing suspended solids at the expense of increased pressure drop or headloss buildup rates, i.e., shorter filter runs. Larger media have reduced initial headlosses and pressure drop increases during the filter run, but may yield a higher suspended solids concentration in the effluent.

The purpose of pilot studies is to quantify the above experiences for the wastewater under study. The quantitative relationships developed must then be used in conjunction with economic and physical design factors to define the optimum design. The historic trend has been toward coarser sizes of filter media in order to attain higher flow rates without reducing the length of filter runs. In order to assure an adequate effluent quality with coarse media and high filtration rates, chemical treatment of the feed may be required.

Another recent trend in filtration has been the adoption of the multi-media concept. Conventional single medium filters have a fine to coarse gradation in the direction of flow which results from hydraulic gradation during backwash. This type of gradation is not efficient as virtually all of the removal and storage must take place in the upper few inches of the filter with a consequent rapid increase in headloss. A coarse to fine filter gradation is much more efficient as it provides for much greater utilization of bed depth, using the fine media only to remove the finer fraction of the suspended solids.

One method of obtaining a coarse to fine filter gradation is the dual-media filtration concept. This employs the use of a layer of coarse anthracite coal over a layer of fine sand. The sizes of the anthracite and sand are chosen so that the coarser but lighter anthracite (specific gravity 1.6) will remain above the heavier (specific gravity 2.65) but smaller sand during backwash. It is desirable to have the coal as coarse as possible to prevent surface blinding and the sand as fine as possible to promote high degrees of removals. However, the disparity in sizes cannot be too great lest overtopping of the coal by the sand would result. In general, sand sizes much finer than 40 mesh are not utilized because the coal size required to prevent overtopping by sand during backwash would be too small to allow high filtration rates. To ascertain the degree of mixing which will occur during backwashing and its effects on subsequent filter performance, pilot column studies are best utilized.

An extension of the dual media concept is the mixed-media filter. Ordinarily this is a tri-media filter of coal over sand over garnet (specific gravity 4.2). There is evidence that judicious selection of the size of each medium will allow a degree of intermixing of the media, such that a reasonable approximation of continuous coarse to fine gradation is obtained. Figure 9-4 illustrates the conventional single medium filter, a non-mixed dual media filter and an ideal coarse-to-fine filter. Typical designs of multimedia filter beds are given in Table 9-1.

Table 9-1

TYPICAL MULTI-MEDIA DESIGNS (6)

Design No.	Garnet		Sand		Coal	
	Size (Mesh)	Depth (Inches)	Size (Mesh)	Depth (Inches)	Size (Mesh)	Depth (Inches)
1	-40x80	8	-20x40	12	-10x20	22
2	-20x40	3	-10x20	12	-10x16	15
3	-40x80	3	-20x40	9	-10x20	8

Conley and Hsiung (6) present additional information on the design of multi-media filters for a variety of applications.

Several studies have compared the performance of single medium, dual media and multi-media filters (7), (8), (9). These studies indicated that in general the latter two types of filters outperformed single medium filters. Better effluents were obtained at higher flow rates, with longer filter runs.

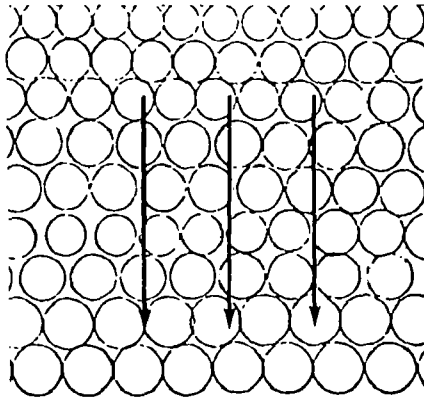
9.2.4 Filter Bed Depth

Although it is known that single medium filters make effective use of only the top few inches of the bed, the historical practice of designing deep beds (24 to 36 in.) has not been abandoned. Dual and multi-media filter depths of the same magnitude are justified because the full filter depth is utilized. In fact, deeper filters of this type may find utility.

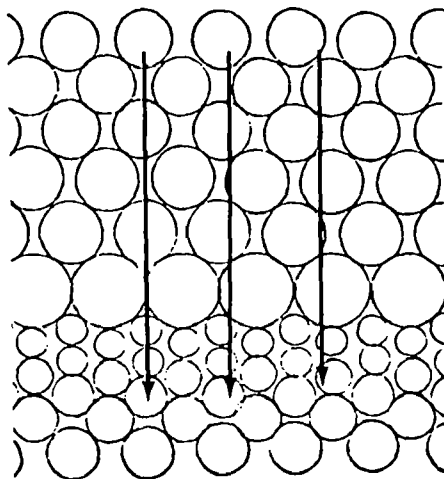
9.2.5 Flow Rates

Traditionally, the design flow rate of most single medium filters was 2 gal/min/ft^2 . Recently these have been raised to 4 gal/min/ft^2 or greater when coarser media were employed along with higher terminal pressure drops. The multi-media filters have been successfully operated at rates up to 8 gal/min/ft^2 on a continuous basis. Rates of this magnitude or higher can be anticipated for design in the future. Of course, pilot studies are essential in determining the design rates of a full-scale installation.

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



Cross-Section Through
Dual-Media Bed
Coarse Coal Above
Fine Sand



Cross-Section Through
Ideal Filter
Uniformly Graded From
Coarse to Fine
From Top to Bottom

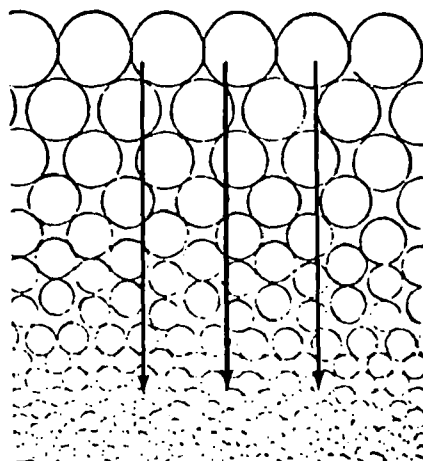


Figure 9-4 MEDIA COMPARISONS

9.2.6 Filter Size, Layout and Housing

Both gravity and pressure filter systems of standard sizes are provided by various filter manufacturers. Specific size details can be obtained from manufacturers' catalogues. Gravity filters are usually constructed on site of concrete. Common feedwater and backwash water supply header pipes generally travel down a central pipe gallery (see Figure 9-1) between rows of filter elements. The gravity filters are usually enclosed in a filter building with storage of finished water in the basement area. All pumps, motors, controls, etc., are housed in the filter building.

Pressure units are more individualized than gravity filters. They are fabricated of steel and are cylindrical in shape. Structural design is according to standard pressure codes. Vessels up to 10 feet in diameter and 60 feet in length are available. It is essential that a manhole be incorporated in the design to allow for maintenance. The filter should be designed with a means for hydraulic removal of all the filter media. Sight glasses for observation of the bed should also be incorporated in the design.

9.2.7 Filter Cleaning Systems

Termination of a filter run takes place when either of these events occurs:

- a. The effluent does not meet the quality criteria
- b. The bed pressure drop is excessive

Either event indicates that the filter is excessively dirty and must be cleaned. The importance of an effective cleaning system on the filter system performance has been reviewed in several recent publications (10), (11). If effective cleaning is not obtained, short filter runs and poor effluent quality will result. Indeed in many respects, improper cleaning sets up self perpetuating operational difficulties such as mud balls, filter media slime coating and media cracking (5).

Backwashing the filter bed by flow reversal is the major, and in many cases, the only method used to clean the bed. The sources of backwash water may include feedwater, filter effluent or some other effluent downstream of the filter. It is virtually impossible to predict the specific conditions required to insure cleaning of the bed by this technique. Experience indicates that for a single medium filter provision should be made to backwash the bed for a 10- to 15-minute period at a rate which will insure fluidization of all the media. The optimum requirements can only be determined by the plant operator once the plant is in service. The plant operator should be instructed that excessive expansion of the bed is just as bad as insufficient expansion. Optimum expansion varies with the size of the medium, type of floc and penetration of floc.

The most important aspect of the design of the backwash system is to insure uniform distribution of the wash water over the entire cross section of the filter area. Although the function of the underdrain system of the filter is both to collect the filtrate and distribute wash water, the latter function controls the design.

A variety of underdrain systems are available for use. The traditional system employs several layers of graded gravel under the filter bed with a lateral header system positioned on the

filter floor. The lateral header system is equipped with orifices and provides the preliminary distribution of the wash water. The final distribution is accomplished as the water moves upward through the gravel. Culp & Culp (12) provide rules for the design of a lateral header system.

Several commercial systems are available which employ patented false bottom distributors in place of the lateral header system. The gravel is then placed on top of the false bottom. Diagrams of Leopold Block and Wheeler Filter Bottom systems are illustrated in Figure 9-5.

Some systems have been devised with the concept of eliminating the gravel layers. One system utilizes porous plates over a false bottom, as shown in Figure 9-5. Another system also utilizes a false bottom with strainers on 12" centers. The strainers are nozzles fabricated of plastic and metal with many small openings.

Table 9-2 presents the usual specifications for gravel, as well as a new design suggested by Baylis (5).

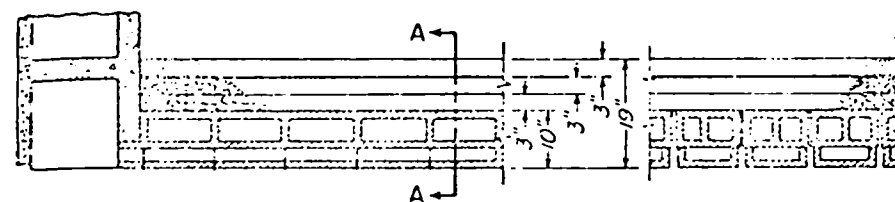
Table 9-2

FILTER GRAVEL DESIGN

Standard		Baylis	
Depth	Size	Depth	Size
2-1/2"	1/12"-1/8"	5"	1" - 2"
3-1/2"	1/8"-1/4"	2"	1/2"-1"
3-1/2"	1/4"-1/2"	2"	1/4"-1/2"
2"	1/2"-3/4"	4"	1/8"-1/4"
4"	3/4"-1-1/2"	2"	1/4"-1/2"
6"	1-1/2"-3-1/2"	2"	1/2"-1"
		4"	1" - 2"

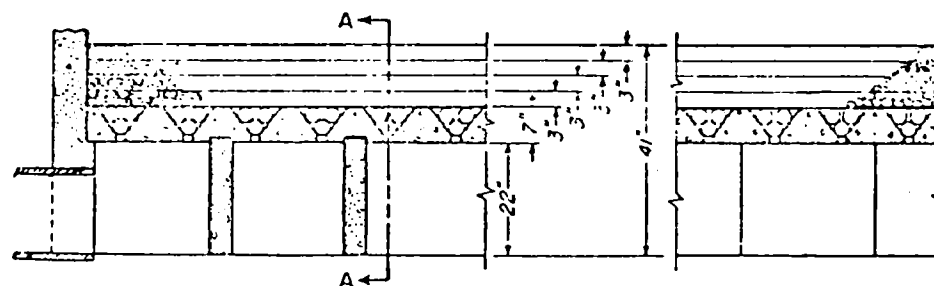
It was found that with the old design the upper layers of gravel could fluidize if excessive backwash rates were used. The Baylis design, which places a final heavy layer of gravel over the finer gravel, prevents this fluidization.

After moving upward through the expanded bed the wash water is conducted out of the filter by wash water troughs. In a pressure filter the inlet baffle serves as the wash water collector. In a gravity filter separate troughs are used because of the large area which must be serviced. In order to prevent non-uniform flow of wash water, the maximum lip to lip distance between troughs cannot exceed six feet. The troughs are usually rectangular channels. A convenient width is assumed and the depth computed by a momentum analysis of the backwater curve in the trough. Figure 9-6 is a nomogram for this solution. For example,



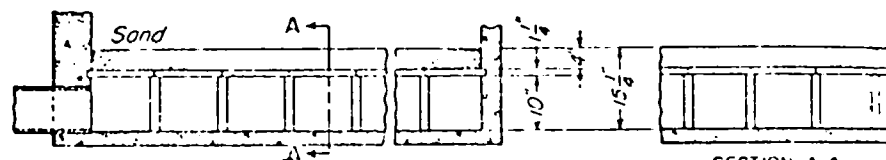
SECTION A-A

A Leopold-block underdrain, with glazed-tile blocks furnishing passages to the water, instead of using lateral pipes.



SECTION A-A

The Wheeler-filter bottom, consisting of solid, inverted, truncated pyramids with water connections at the apex of each pyramid, and the pockets filled with cement or glazed earthenware spheres.



SECTION A-A

Porous-plate filter bottoms.

Figure 9-5 FILTER BOTTOM AND UNDERDRAIN SYSTEMS

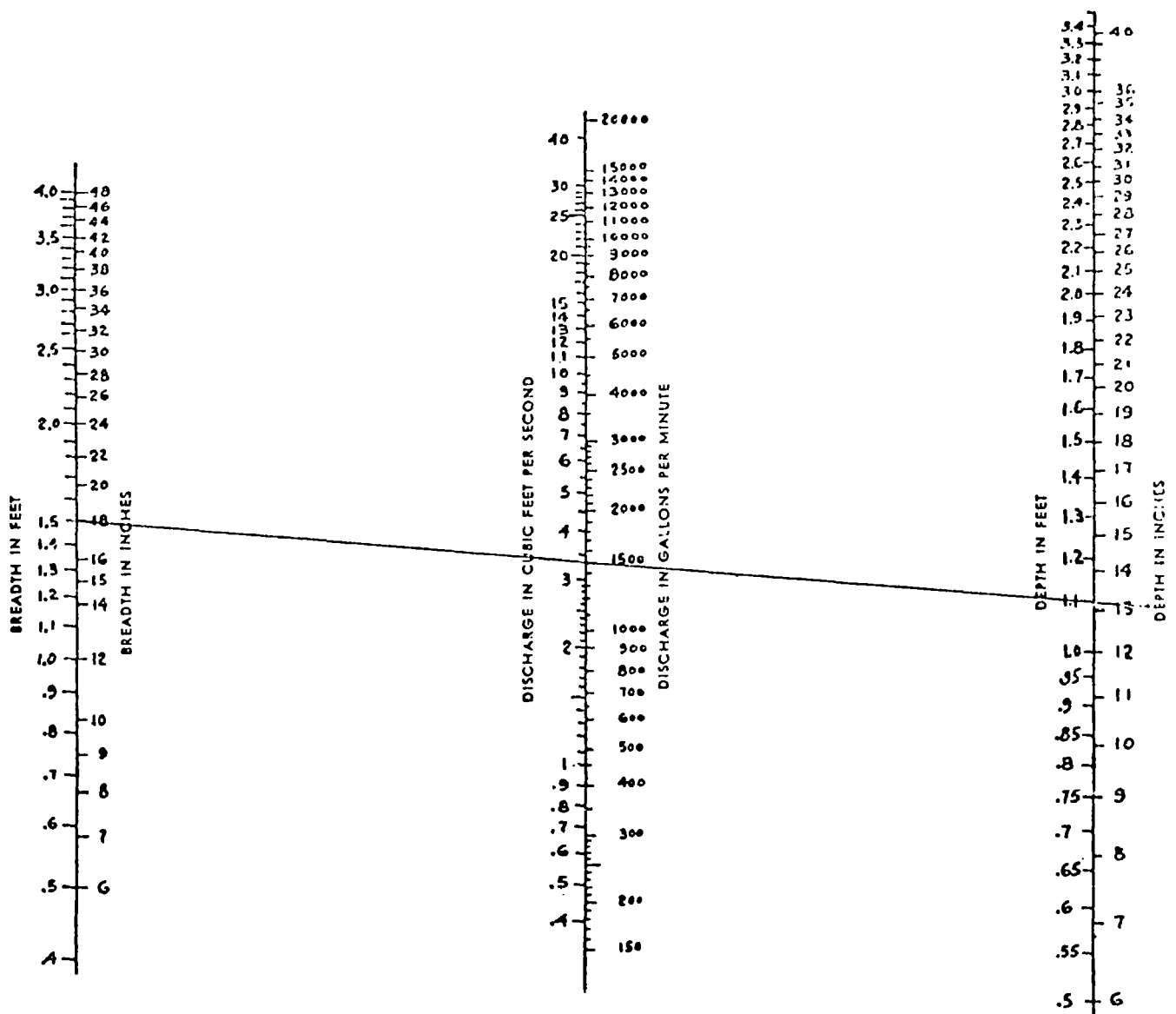


Figure 9-6
NOMOGRAM FOR SELECTING WASH TROUGH DIMENSIONS

an 18 inch wide trough carrying a flow of $3.33 \text{ ft}^3/\text{sec}$ will have a maximum water depth of 13.2 inches. The side wall depth should be increased by 2 to 3 inches to allow for freeboard.

Maintaining a clean bed by backwash alone has been found to be difficult when wastewater, even after biological and/or chemical treatment, is the filter feed. Even periodic chlorination during backwashing has not always been successful in preventing slimes. Consequently, either surface wash or air scour should be employed in waste treatment filters. Multi- and dual-media filters allow for greater floc penetration of the bed than single medium filters. The use of surface wash and/or air scour is mandatory under these circumstances. In essence, the function of air scour and surface wash is to loosen the accumulated deposits in the filter. The normal backwash then flushes the deposits away.

A surface wash apparatus is illustrated in Figure 9-7. Either fixed nozzles or rotating pipes fitted with nozzles are placed about 1 to 2 inches above the top of the bed. While the surface wash is on, the backwash expansion is set at a lower rate than after the surface wash is terminated. Surface wash water is supplied at 50 to 100 psi at rates approximating 1 to 3 gal/min/ft² of bed.

Air scour is accomplished by injecting air into the underdrain system prior to initiating water backwash. The following procedure has been recommended (12):

- a. Stop influent and lower the water level to a few inches above bed.
- b. Apply air alone at $2\text{-}5 \text{ ft}^3/\text{min}/\text{ft}^2$ for 3-10 min.
- c. Apply water backwash at $2\text{-}5 \text{ gpm}/\text{ft}^2$ with air on until water is within one foot of wash water trough.
- d. Shut air off.
- e. Continue water backwash at normal rate for usual period of time.
- f. Apply backwash for 1-2 min. at a rate required to insure hydraulic classification of the filter media.

Air backwash has the disadvantage of increased possibility for coal media losses. Although small losses repeatedly occur due to air bubble attachment during normal backwashing cycles, the danger of massive losses exists if air is applied during trough overflow periods. Control is especially difficult in pressure filters as it is hard to observe the bed during backwash.

9.2.8 Filter Control Systems

Recent improvements and advances have been made in the systems which are used to control the operation of filters. As a result, where once a filter required a manual operation, it is now possible to have a completely automated filtration plant. Not only have these advances removed much of the drudgery from filter operation, but they have given the plant operator powerful techniques for improving filter performance.

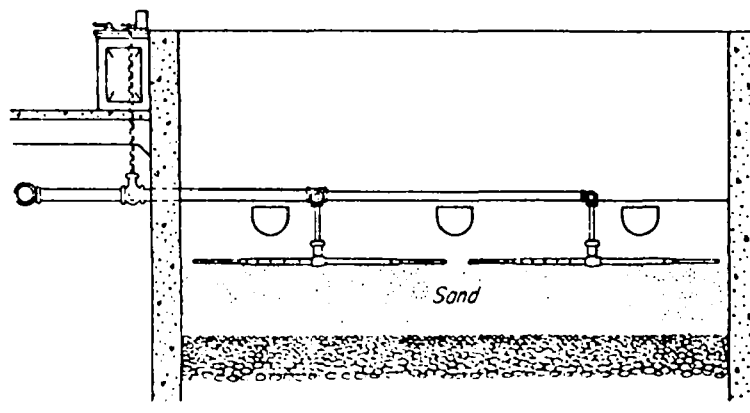


Figure 9-7
PALMER FILTER BED AGITATOR

Perhaps the most important of the recent advances is the use of automatic turbidimeters to continuously monitor the filter feed and product. This allows the operator to anticipate difficulties from changes in feed quality, and rapidly remedy process failures. In addition, these devices allow the operator to rapidly evaluate the effects of changes in process variables and provide a continuous record of plant performance. All turbidimeters operate on the principle of measurement of scattered or transmitted light. A variety of commercial instruments are available.

Automatic backwash, initiated either by terminal pressure drop across the bed or by excess turbidity in the effluent, should be considered for all new plants. The cost of these systems may be justified by a reduction in labor. Automatic backwash systems are available from a variety of manufacturers. It is important that these systems be equipped with delay mechanisms so that the wash water reservoir capacity is not exceeded if the wash cycles of several filters overlap.

The automatic backwash system referred to above is essentially an electronics package which operates valves, pumps, etc., by remote control. Two other automatic backwash systems based on a different concept are commercially available. An automatic gravity filter is depicted in Figure 9-8. No data are available for waste treatment application of these systems.

A traveling backwash filter employs an 11-inch depth of sand supported on porous plates. The bed is divided into many sections 8 inches wide. A traveling backwash assembly moves from section to section as required to clean each section. Thus most of the filter is in operation with only a small section being washed at any time.

In the past, one type of flow control system was utilized in most filters. This system provides for constant flow with a constant water depth over the filter. Under this condition, constant flow is maintained by varying the headloss downstream of the filter so that the total headloss in the filter and downstream is constant. Headloss in the downstream line is automatically adjusted by a throttle valve connected to a preset counterweight. Usually a venturi with a variable opening diaphragm serves as the rate controller. Frequently maintenance of these rate controllers is troublesome.

An alternate system which does not require a rate controller to achieve constant flow has recently been described (13). A weir in the inlet channel to the clear well provides a constant back pressure on the filter. Flow into the filter box is over an inlet weir with the crest set quite high above the bed surface. As headloss develops in the filter, the depth of flow above the filter increases to maintain the filter flow constant. A disadvantage of this method is the capital cost associated with building the filter box walls several feet higher than required where a rate controller is used.

Another control concept is declining-rate filtration. This method is applicable only to medium or large scale plants which utilize multiple filters. With this method the flow rate through the filter is allowed to decline as the filter clogs. The filters are staggered in degree of clogging so that the total production of the plant is constant. This procedure is claimed to produce better effluent quality and longer filter runs because flow slows as the filter clogs. Thus, the rate of headloss increase decreases and the probability of floc breakthrough is lessened.

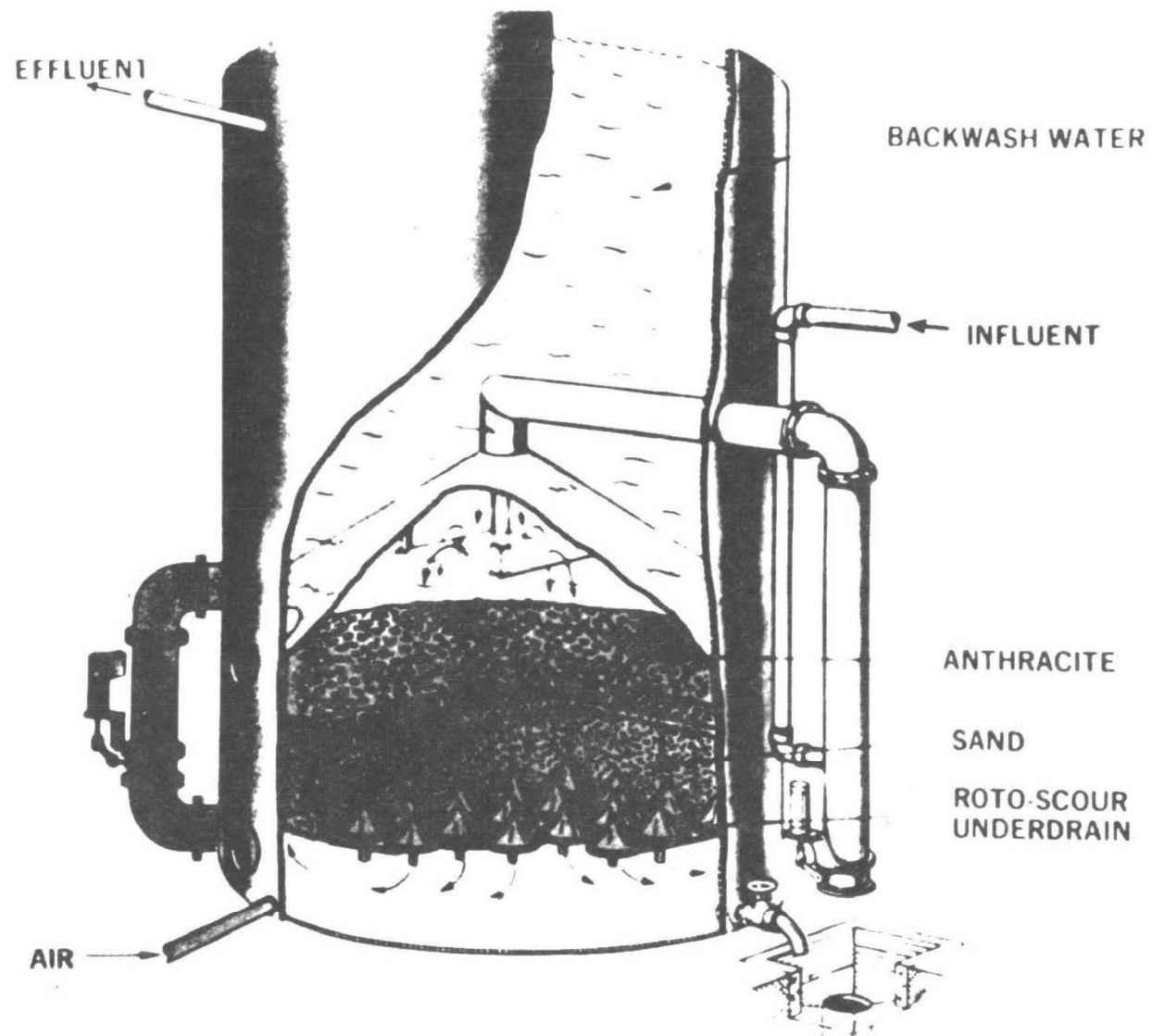


Figure 9-8 AUTOMATIC GRAVITY FILTER, SINGLE COMPARTMENT
(Courtesy of Graver)

A simple method of flow control which can be used with pressure filters is to pump at a constant rate to each filter with a positive displacement pump. With this system, the pump rate sets the flow rate and the pump discharge pressure rises as the bed clogs.

9.2.9 Chemical Pretreatment

Chemical pretreatment ahead of the filter is designed to:

- a. Coagulate suspended solids to make them more amenable to sedimentation and/or filtration.
- b. React with soluble components which must be removed to form insoluble precipitates for removal by sedimentation and/or filtration.
- c. Adjust the strength of the floc to control the degree of penetration of solids into the filter.

In order to ascertain optimum chemical doses to accomplish any of these objectives, pilot studies are required.

While coagulants aid in removal of colloids, the floc formed may be relatively weak. The major use of polymeric materials is to strengthen floc so that it will not penetrate through the filter. The floc should not be excessively strengthened or it will not penetrate beyond the filter surface, producing excessive pressure drop. A good rule to follow is that floc breakthrough should coincide with the achievement of terminal headloss.

Chemicals are usually applied prior to sedimentation to remove the bulk of the solids before the filter. Recent practice has employed further chemical addition just prior to filtration to enhance the filterability of the feedwater solids.

9.3 Summary of Results of Filtration Studies

Rapid sand type filters have been used in three kinds of waste treatment systems: direct filtration of secondary effluent, filtration of secondary effluent after chemical treatment and filtration of raw or primary wastewater after coagulation and sedimentation. Representative results from field studies in each of these areas will be discussed below.

9.3.1 Filtration of Secondary Effluent

Early work on filtration of secondary effluent took place in Europe. Truesdale and Birkbeck (14) reported on tests run between October, 1949, and May, 1950, at the Luton Sewage Works. Beds of sand 2 feet deep, ranging in size from 0.9 mm to 1.7 mm, exhibited 72 to 91 percent removal of suspended solids and 52 to 70 percent removal of BOD. Flow rates ranged from 1.33 to 3.3 Imp. gal/min/ft². Air-scour-aided backwash was used once per day to clean the bed. Backwash flow rate was 11.6 Imp. gal/min/ft².

Naylor, Evans and Dunscombe (15) later reviewed 15 years of studies of tertiary treatment at Luton. A 3-foot deep bed of -10 to +18 mesh sand consistently provided an effluent of 4 to 6 mg/l suspended solids at flow rates of 3.3 Imp. gal/min/ft². It was found best to wash the beds every 12 hours.

In the U.S., most direct filtration work has been with activated sludge feed. At the Hyperion Plant in Los Angeles, sand of 0.95 mm effective size was used in a shallow bed (11 inches deep) traveling backwash filter. This study lasted for six months during which time 46 percent suspended solids removal and 57 percent BOD removal were obtained. Filtration rate was 2 gal/min/ft². Difficulty was encountered in cleaning the filters and performance gradually deteriorated during the study. Use of a finer sand (0.45 mm effective size) in an attempt to yield a better effluent was a failure due to very rapid clogging of the filter (16).

Much greater success utilizing the traveling backwash filter for activated sludge effluent treatment was obtained by Lynam (17) in Chicago. The effective size of sand used in this study was 0.58 mm. Suspended solids removal of 70 percent and BOD removal of 80 percent were obtained at flow rates of 2 to 6 gal/min/ft². Terminal headloss was quite low (11 inches of water.) The range of flows studied exhibited no significant difference in terms of suspended solids removal.

A study of filtration of activated sludge effluent was presented by Tchobanoglous and Eliassen (18). They found that flow rate (2 to 10 gal/min/ft²) had little effect on performance, but effective size of sand in the range of 0.4 mm to 1.2 mm had a very significant effect. It was also determined that virtually all suspended solids removal took place in the upper six inches of the bed. It was concluded that the floc strength of activated sludge is quite high on the basis of the low degree of penetration even with very coarse media and high flow rates. Poor removal of suspended solids (10 to 40 percent) was obtained in this study, with filter depths greater than 6 inches producing no additional benefits. It was found that a bimodal distribution of particle sizes existed for the activated sludge effluent. Apparently, only the larger solids could be removed by the filters.

Culp and Culp (12) reviewed the work on plain filtration of secondary effluent with both single medium and multimedia filters. They concluded that, with either type of filter, better results would be obtained as the degree of self flocculation of the sludge increased. Thus, a high-rate activated sludge effluent which contains much colloidal material should filter poorly, while an extended aeration effluent should filter well. Multi-media filters exhibit a marked superiority for filtration of activated sludge effluent because of the high volume of floc storage available in the upper bed and the polishing effect of the small media. They indicated the expected performance of multi-media filters for plain filtration of secondary effluents, as shown in Table 9-3.

Table 9-3

EXPECTED EFFLUENT SUSPENDED SOLIDS FROM MULTI-MEDIA
FILTRATION OF SECONDARY EFFLUENT

Effluent Type	Effluent S.S. mg/l
High Rate Trickling Filter	10 - 20
2-Stage Trickling Filter	6 - 15
Contact Stabilization	6 - 15
Conventional Activated Sludge	3 - 10
Extended Aeration	1 - 5

9.3.2 Filtration of Chemically-Treated Secondary Effluent

Treatment of secondary effluent by chemical coagulation, sedimentation and filtration has been conducted at a number of installations. The purposes of this treatment procedure have often been twofold, suspended solids and phosphorus removal. Unfortunately, in many cases, performance results have been reported for complete systems, not each individual process. Thus, analysis of the filter performance is not possible. In most of these installations, the filter is viewed essentially as a polishing device to capture solids which escape the sedimentation tank. As Culp and Culp (12) have indicated, this philosophy may be wrong, as modern filters can absorb much greater solids loads than older designs.

In conjunction with the studies at Chicago (17), coagulation with alum followed by filtration was evaluated. It was found that the alum treatment had little effect on the effluent quality. It is probable that this result is due to the weakness of chemical floc compared to activated sludge floc.

An advanced waste treatment plant has been used to renovate step aeration activated sludge effluent in Nassau County, New York. Alum at 200 mg/l is used to coagulate the wastewater prior to sedimentation and filtration. The filters are dual-media containing 30 inches of 0.9 mm coal over 6 inches of 0.35 mm sand. With the addition of 0.5 mg/l of an anionic polymer, effluent turbidity is maintained below 0.4 JTU. Run length varies from 8 to 24 hours depending on the solids load from the clarifier.

At Lebanon, Ohio, treatment of the activated sludge effluent with lime has been investigated (19). The lime dose averaged 300 mg/l. Dual-media filters, consisting of 18 inches of 0.75 mm coal over 6 inches of 0.45 mm sand, followed clarification. The filters were operated at 2 gal/min/ft², and were backwashed when the headloss reached 9 ft of water. Influent to the filters ranged from 13 to 36 mg/l of suspended solids (turbidity 4 to 10 JTU.) Filter effluent ranged from 0.07 to 0.14 JTU.

At Lake Tahoe, lime is used to coagulate activated sludge effluent prior to sedimentation and filtration (12). Multi-media beds 3 feet deep are utilized at an average flow rate of 5 gal/min/ft². Filter runs have varied from 4 to 60 hours. Polymers as well as alum have been used to strengthen the floc. Normally, runs are terminated at headlosses of 8 ft of water. Effluent turbidities are typically reduced to 0.3 JTU with correspondingly low values for other parameters.

At the Environmental Protection Agency pilot plant at Washington, D.C., mineral addition to the final phase of a step aeration activated sludge is practiced for phosphorus removal. After sedimentation the effluent is filtered through parallel filters at 2.4 gal/min/ft². One filter is a dual-media, while the other is a multi-media. It has been found that the multi-media filter removes 5 to 10 percent more suspended solids than the dual-media filter. Filter runs have been in the range of 24 to 32 hours. Typically, the filters reduce secondary effluent suspended solids from 33 mg/l to 8 mg/l (20).

9.3.3 Filtration Following Chemical Treatment of Primary or Raw Wastewater

Clarification of raw wastewater followed by carbon adsorption is just emerging as a viable treatment technology. This system employs filtration as part of the solids separation system. Although several pilot installations employing this concept are in operation, filtration has not been closely studied; thus data are sparse.

At the EPA pilot plant in Washington, D.C., two-stage lime treatment of raw sewage followed by sedimentation, filtration and granular carbon adsorption is being studied (21). Dual-media filters (18 inches of 0.9 mm coal over 6 inches of 0.45 mm sand) are employed. Cleaning is initiated at a headloss of 9 feet of water. Cleaning is performed automatically with a surface wash rate of 3 gal/min/ft² and an upflow rate of 20 gal/min/ft². Run lengths averaged 50 hours during cold weather, but in warm weather the growth of slimes reduced run lengths to less than 12 hours. Prechlorination of the filter feed was employed to restore the run lengths to 50 hours. Filter effluent averaged 4.5 mg/l of suspended solids over a 6-month period, which represented a 70 percent efficiency for the filter. This plant operates on a programmed diurnal flow variation which produces a flow rate variation on the filter from 1.7 to 4.3 gal/min/ft².

A similar system using single-stage lime treatment was run for several months at the EPA installation in Lebanon, Ohio (22). Flow to the filters was 2 gal/min/ft². Suspended solids in the filter effluent averaged 10 mg/l, which represented a 67 percent removal efficiency.

9.4 New Filtration Systems

The rapid sand type filtration system previously discussed has been a downflow, batch, static bed system. Consequently, it suffers from a variety of process deficiencies including:

- a. The need to stop the process periodically to clean the filter medium.
- b. The limited ability to economically handle suspensions containing high concentrations of suspended solids.

During the last decade a number of new filtration systems have been developed which are aimed at overcoming these shortcomings. Several of these will be discussed below.

9.4.1 Upflow Filtration

As indicated previously, a major difficulty with downflow, single-media filtration is that after backwash, the bed is graded fine to coarse in the direction of flow. If filtration is conducted upflow, this difficulty is circumvented. However, once the headloss produced by the upflow exceeds the buoyant weight of the filtration medium, fluidization with consequent loss of filtration efficiency will result. One solution to this problem is to place a restraining grid on or near the top of the filter medium to prevent fluidization. A diagram of an upflow filter is illustrated in Figure 9-9 (courtesy of DeLaval.)

The spacing between bars of the grid must be large enough to allow the bed to expand during backwash, but must be small enough to prevent upward bed movement during filtration. It would seem that these two requirements are directly contradictory; however, arching of the grains takes place between the bars, allowing a reasonably large spacing. Space between the bars is usually in the range of 100 to 150 diameters of the smallest grain size in the beds. During cleaning, air is first introduced to agitate the bed. After the air has broken the arches, backwashing with water is started. Table 9-4 gives a summary of typical design parameters for the upflow filters.

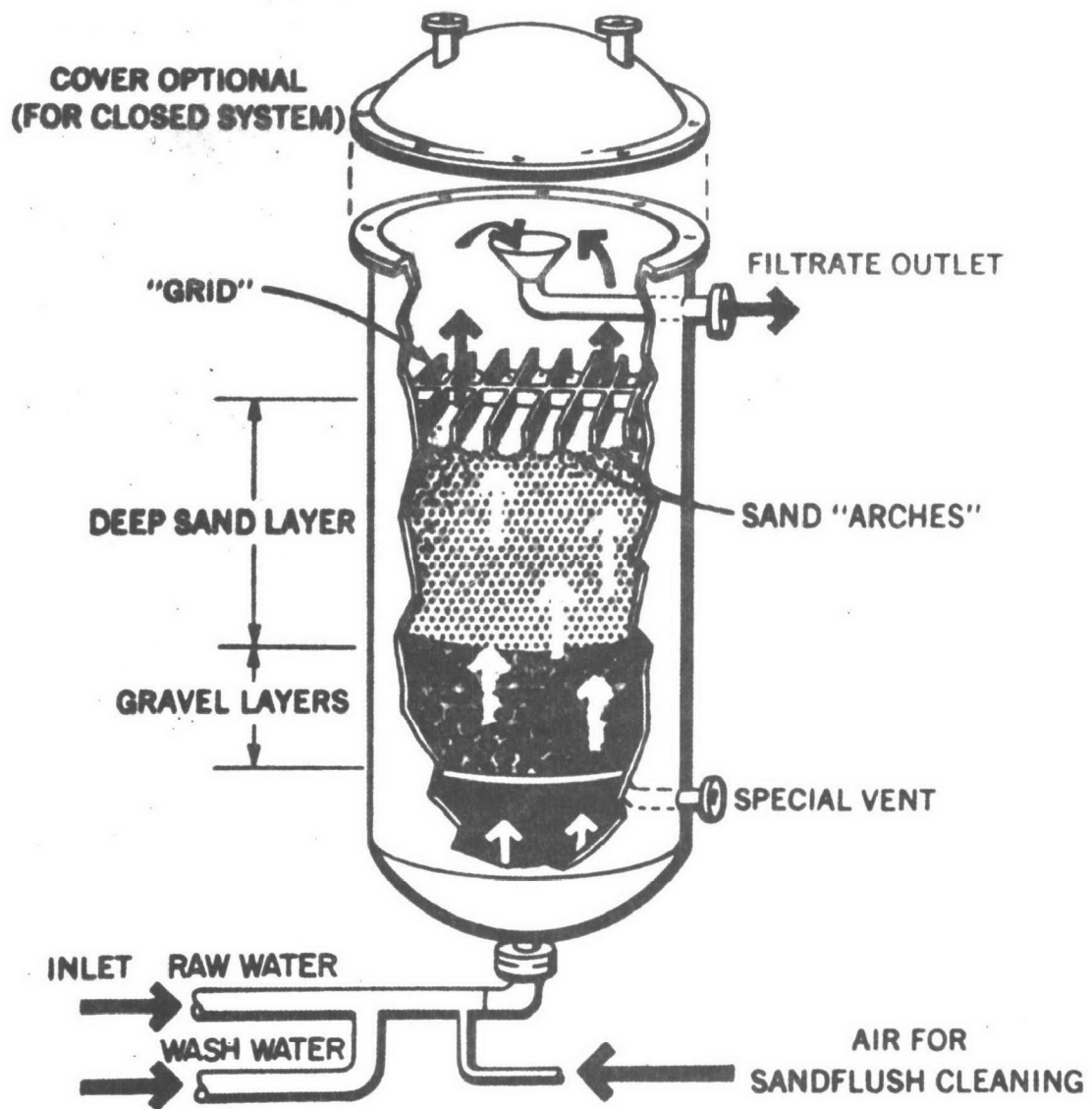


Figure 9-9 CROSS SECTION OF UPFLOW FILTER

Table 9-4

DESIGN PARAMETERS FOR UPFLOW FILTERS

Bed Material:	Sand
Bed Construction:	60 in. 1 - 2 mm
	10 in. 2 - 3 mm
	4 in. 10 - 15 mm
Flow Rate:	2-3 gal/min/sq ft
Backwash Rate:	To achieve minimum 20% expansion
Terminal Headloss:	6 to 20 feet of water

Boby and Alpe (23) reported on the performance of an upflow filter treating secondary effluent at Totor, England. Average suspended solids removal was 85 percent, with the filter effluent below 5 mg/l. These results were equal to or better than those obtained with downflow filtration with the same size bed. It is claimed that this type of filter can absorb higher loads of solids than a conventional filter.

9.4.2 Moving Bed Filter

The basic concept of a moving bed filter is the mechanical movement of the most heavily clogged portion of the medium out of the zone of filtration with virtually no interruption of the filtration process. The potential of such a process is operation at higher flow rates and at much higher solids loadings than conventional systems. Superior cleaning of the filter media should also be possible.

Johns-Manville Corporation has developed a moving bed filtration system. A diagram illustrating the essentials of this system is given in Figure 9-10. Wastewater (A) flows through the inlet pipe where chemicals, if required, are added at (B). The wastewater enters the head tank (C) and then passes through the sand bed (D). The filtered water leaves through the exit screens. When excessive headloss develops, the bed is pushed toward the head tank by pressurizing a chamber separated from the bed by a flexible diaphragm. A mechanical cutter (F) sweeps down over the face of the bed cutting off the top layers. These then fall into the hopper (G) of the head tank. The sludge and sand are removed from the head tank with the aid of an ejector using feedwater. The solids are hydraulically conveyed to the sand washer (H) where filtered water or air and filtered water are used to backwash the sand. Clean sand moves by gravity back to the base of the filter. The spent washwater is sent to a sedimentation tank for removal of the wastewater solids. The operation of the system is automated.

Under an Environmental Protection Agency contract (24), this system was evaluated by the manufacturer for the treatment of raw wastewater, primary effluent and settled and unsettled trickling filter effluents at the Bernards Township Sewage Treatment Plant. Alum was used to precipitate phosphorus, and an anionic polymer was employed to prevent excessive floc penetration. The results, given in Table 9-5, show excellent treatment performance in these situations.

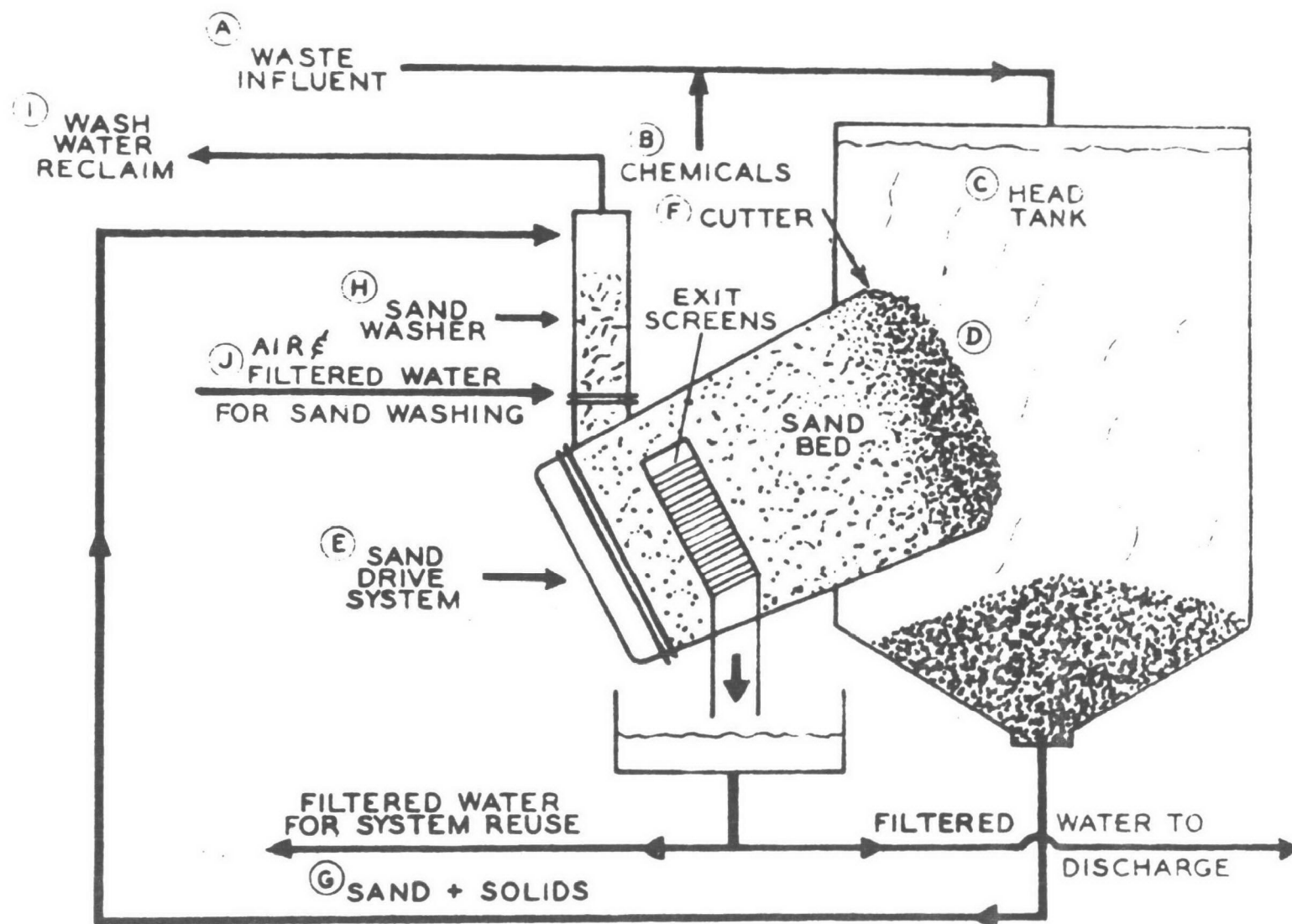


Figure 9-10

SCHEMATIC DRAWING OF THE JOHNS-MANVILLE MOVING BED FILTER

The moving bed filter is being evaluated at full scale (2 MGD) at the Borough of Manville Sewage Treatment Plant, Manville, N.J. At this site, unsettled trickling filter effluent is the feed. This study is being partially funded by an EPA demonstration grant.

Table 9-5

**JOHNS-MANVILLE MOVING BED FILTER EVALUATION AT
BERNARDS TOWNSHIP SEWERAGE AUTHORITY TREATMENT PLANT**

Parameter (mg/l)	Final Effluent w/o Chlorination			Unsettled Trickling Filter Effluent			Primary Effluent			Raw Wastewater		
	In	Out	%	In	Out	%	In	Out	%	In	Out	%
Total P	9.37	0.51	95	19.1	0.99	95	14.6	1.13	93	21.5	2.16	91
Filterable P	8.03	0.11	99	14.9	0.62	96	13.2	0.58	96	18.6	0.79	96
Ortho P	7.80	0.10	99	12.4	0.53	96	9.8	0.38	96	13.2	0.57	95
BOD ₅	65	12	80	55	3.8	93	67	12	82	115	19	84
Suspended Solids	50	15	70	86	7.1	91	77	11	87	156	27	83
Turbidity (JTU)	33	7	79	39	3.4	91	53	3.7	93	123	16.7	87

Alum: 200 mg/l (commercial grade)

Polyelectrolyte: 0.5 mg/l anionic

The moving bed filter system is currently available in modules of 2-bed and 4-bed configurations. These modules can be used singly or in combination to accommodate large flow requirements as necessary.

- The flow rate through any unit or combination of units is dependent on the quality of the incoming liquid and the discharge requirements. Flow rates up to 7.0 gal/min/ft² of exposed filter area are possible.
- Filter bed dimensions - 48-inch face diameter, 60-inch length from the face of the unit to the center line of exit screen.

c. Module dimensions -

	<u>2-Bed Unit</u>	<u>4-Bed Unit</u>
Plan length	20'	20'
Plan width	7'	14'
Height	17'	17'
Weight, empty	22,000 lb	44,000 lb
operating	78,000 lb	156,000 lb
Plan floor space, module system	140 ft ² 350 ft ²	280 ft ² 720 ft ²

d. Power requirements -

Function	Two-Bed System		Four-Bed System	
	Conn. HP	Op. HP	Conn. HP	Op. HP
Diaphragm or Sand-bed Movement	3.0	3.0	6.0	6.0
Cutter	3.0	0.75	3.0	1.5
Sand Cleaning System	3.75	3.75	7.5	7.5
Screen Wash	7.5	0.5	7.5	1.0
Chemical Pump and Mixer	1.0	0.5	1.0	0.75
Air Compressor	<u>1.0</u>	<u>0.5</u>	<u>1.0</u>	<u>0.75</u>
	19.25	9.0	26.0	17.5

e. Filter media - Hard sharp-grained quartz sand of filter grade quality (effective size 0.6 to 0.8 mm; uniformity coefficient 1.5)

f. Sand drive rate -

Linear - 12 in./hour - maximum
Volume - 25 ft³/hour - maximum
Pump pressure - 75 psi - average
 150 psi - maximum

g. Head tank - 30 minute detention time

h. Sludge settling tank - 30 minute detention time

i. Controls -

Low level and high level off-on
Differential pressure to actuate bed movement and cutter
Effluent turbidity to adjust chemical feed

9.4.3 Radial Flow Moving Bed Filter

Recently, Dravo Corporation has introduced a radial flow moving bed filter to the American market. At present, most applications of this system have been for industrial waste treatment. In addition to the moving bed concept, the main feature of this system is the radial flow concept. This geometric configuration provides more filter area per unit volume than downflow or upflow systems. As the liquid flows radially from the central core, it slows down, providing increased opportunity for solids removal.

9.4.4 Radial Flow-External Wash-Filter

The Hydromation Corporation has developed a new concept in filtration. A diagram of this filter is illustrated in Figure 9-11. It is a batch-type radial flow filter with external media wash. When a filter run is terminated, the media is pumped out of the filter and upward into a scrubber. The flow velocity in the scrubber is 20 ft/sec, which produces a very high degree of turbulence, assuring good cleaning of the media. The clean media circulates around the backwash loop to the radial flow bed.

This filtration system utilizes a polymer resin as the filter medium. It is claimed to have superior dirt holding capacity compared to natural media. Because of the superior dirt holding character and special cleaning system, flow rates of 10 to 20 gal/min/ft² have been purported.

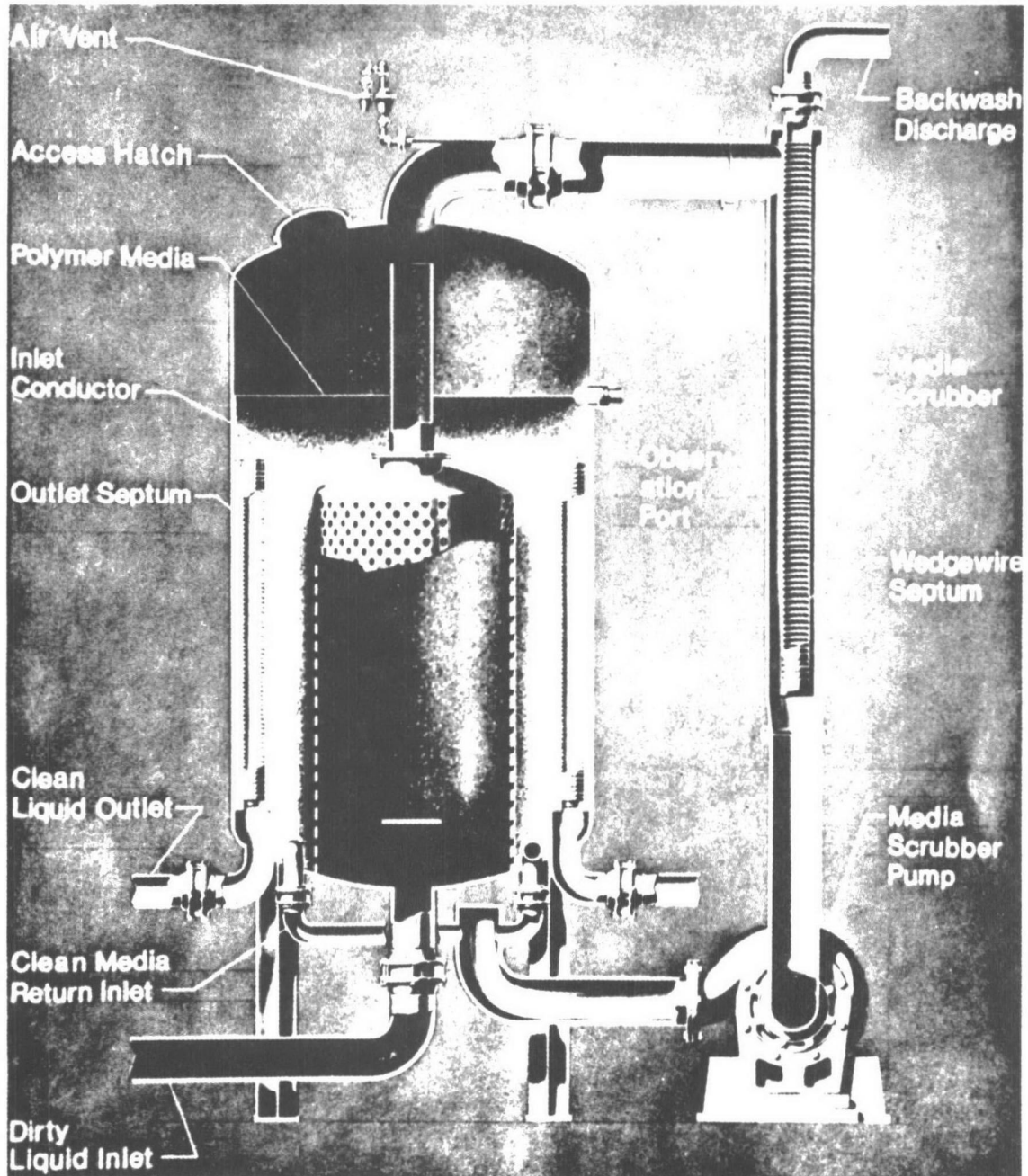


Figure 9-11 HYDROMATION IN-DEPTH FILTER

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