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COST AND PERFORMANCE ESTIMATES FOR TERTIARY WASTEWATER TREATING PROCESSES

ADVANCED WASTE TREATMENT RESEARCH LABORATORY - IX

U.S. DEPARTMENT OF THE INTERIOR
FEDERAL WATER POLLUTION CONTROL ADMINISTRATION
OHIO BASIN REGION
Cincinnati, Ohio

COST AND PERFORMANCE ESTIMATES FOR TERTIARY WASTEWATER TREATING PROCESSES

by
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FOREWORD

In its assigned function as the Nation's principal natural resource agency, the United States Department of the Interior bears a special obligation to ensure that our expendable resources are conserved, that renewable resources are managed to produce optimum yields, and that all resources contribute their full measure to the progress, prosperity, and security of America -- now and in the future.

This series of reports has been established to present the results of intramural and contract research studies carried out under the guidance of the technical staff of the FWPCA Robert A. Taft Water Research Center for the purpose of developing new or improved wastewater treatment methods. Included is work conducted under cooperative and contractual agreements with Federal, state, and local agencies, research institutions, and industrial organizations. The reports are published essentially as submitted by the investigators. The ideas and conclusions presented are, therefore, those of the investigators and not necessarily those of the FWPCA.

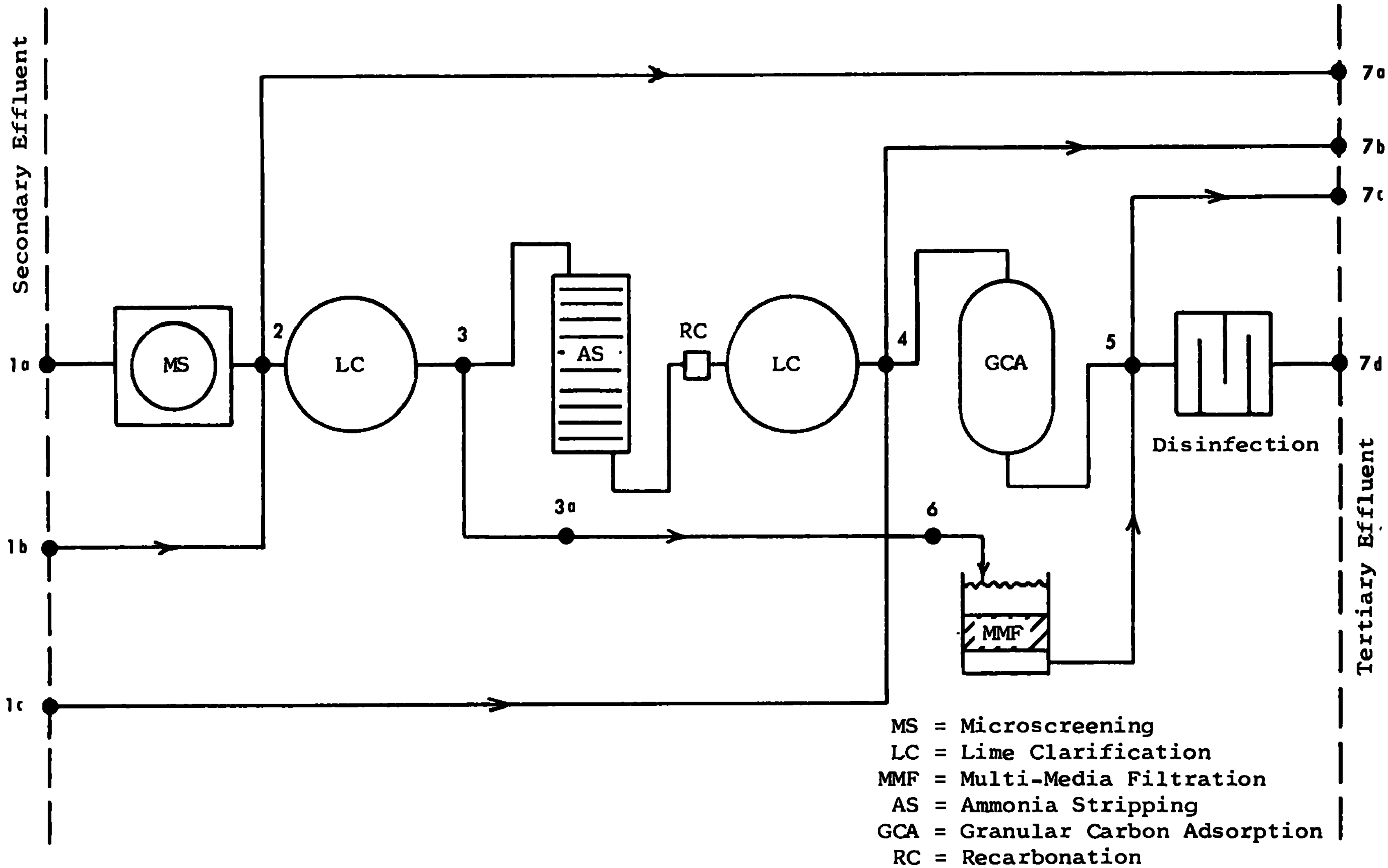
Reports in this series will be distributed as supplies permit. Requests should be sent to the Office of Information, Ohio Basin Region, Federal Water Pollution Control Administration, 4676 Columbia Parkway, Cincinnati, Ohio 45226.

COST AND PERFORMANCE ESTIMATES FOR TERTIARY WASTEWATER TREATING PROCESSES

Robert Smith and Walter F. McMichael
Treatment Optimization Research Program

This report contains generalized estimates of both performance and cost for wastewater treatment processes which can be used downstream of the activated sludge process to reduce the pollution load on the receiving stream. Cost and performance estimates given are believed to be the most valid and up-to-date information now available. No attempt has been made to treat every process, process modification, or process group proposed for tertiary treatment. Processes treated reflect only the current thinking in this technological area. Processes and groups of processes believed to be leading candidates for use downstream of secondary treatment are shown in Figure 1. The group of processes selected for use by the process designer will depend on the water quality requirements at the receiving stream. For example, if only partial removal of organic contaminants such as BOD, COD, or TOC is required, microscreening or rapid filtration can be used to remove about 70% of the suspended solids which represent a significant portion of the organic contaminant.

Data from various sources on the fraction of 5-day BOD associated with suspended solids are shown in Table 1. Similar data for COD are given in Table 2. Obviously, this fraction is strongly dependent on the operating mode of the aerator as well as the efficiency of the final settler in preventing suspended solids



WASTEWATER TREATMENT PROCESSES FOR USE DOWNSTREAM OF SECONDARY TREATMENT

FIGURE 1

Table 1

MEASUREMENTS ON THE FORM OF ORGANIC SPECIES IN ACTIVATED SLUDGE PROCESS EFFLUENT

	F/T BOD[β]	TBOD[α]	Delta BOD/Delta SS[γ]
Hyperion (3 weeks)			
1. 6 hr detention, 2000 mg/l	.57	4.4	.35
2. 6 hr detention, 3000 mg/l	.72	5.0	.18
3. 5 hr detention, 3500 mg/l	.61	8.0	.214
4. 4 hr detention, 3000 mg/l	.53	9.0	.27
5. 4 hr detention, 2000 mg/l	.25	12.2	.35
Washington, D. C. Blue Plains Plant (2 months)			
2.5 hr detention, 444 mg/l	.39	45.0	.70
26th Ward Plant, New York (6 years)			
2.4 hr detention, 260 mg/l	.44	36.0	.61
Tallmans Island Plant (6 years)			
4.0 hr detention, 773 mg/l	.44	19.0	.40
Hunts Point Plant, New York (6 years)			
2.8 hr aeration, 750 mg/l	.40	18.5	.55
Rockaway Plant, New York (6 years)			
1.8 hr aeration, 480 mg/l	.43	33.0	.55
Jamaica Plant, New York (6 years)			
1.6 hr aeration, 930 mg/l	.52	35.0	.41

α TBOD = Total 5-Day BOD (Dissolved + Particulate)
β F/T = Filtrate 5-Day BOD/TBOD
γ (TBOD - Filtrate BOD)/Suspended Solids

Table 2

FILTERED AND UNFILTERED CHEMICAL OXYGEN DEMAND MEASUREMENTS
ON
ACTIVATED SLUDGE EFFLUENT

Hyperion Treatment Plant Los Angeles, California	VSS mg/l	TCOD mg/l	DCOD mg/l	SCOD/TCOD	ΔCOD/VSS
1. 6 hr. detention, 2000 mg/l MLSS	4.4	29.8	26.1	.12	.84
2. 6 hr. detention, 3000 mg/l MLSS	6.0	33.8	27.4	.19	1.1
3. 5 hr. detention, 3500 mg/l MLSS	12.5	40.0	28.4	.29	.93
4. 4 hr. detention, 3000 mg/l MLSS	11.6	42.0	30.4	.28	1.0
5. 4 hr. detention, 2000 mg/l MLSS	22.8	82.4	41.5	.50	1.8
6. 4 hr. detention, 1000 mg/l MLSS	16.6	91.4	49.5	.46	2.5
7. 6 hr. detention, 1000 mg/l MLSS	13.0	56.6	43.3	.24	1.0

Pomona Treatment Plant Pomona, California	SS mg/l	TCOD mg/l	DCOD mg/l	SCOD/TCOD	ΔCOD/SS
1. Mean Cell Residence Time = 4.9 days	8.0	53.0	44.0	.17	1.1
2. Mean Cell Residence Time = 4.8 days	20.0	82.0	54.0	.34	1.4
3. Mean Cell Residence Time = 9.4 days	11.0	49.0	36.0	.27	1.2
4. Mean Cell Residence Time = 9.3 days	7.6	39.0	27.0	.31	1.6

VSS = Volatile Suspended Solids
SS = Suspended Solids
TCOD = Unfiltered COD
ΔCOD = Filtered COD
ΔCOD = TCOD - DCOD

from escaping in the effluent stream. For a reasonably well designed and operated activated sludge plant, the concentration of various contaminants that might be expected in the effluent stream are shown in Table 3.

Based on the values shown in Tables 1 and 2, it has been assumed that 60% of the 5-day BOD is in the form of particulate. COD and TOC were assumed to be 70% dissolved and 30% particulate. Microscreening or rapid sand filtration can, therefore, be expected to remove about 42% of the 5-day BOD and 21% of the COD and TOC. Other solids removing processes, such as lime clarification, multi-media filtration, and granular carbon adsorption, will remove a greater fraction of the suspended solids. Some small fraction of the dissolved organic contaminants might be removed, but for a first approximation this appears to be negligible. A large fraction of the dissolved organic species is removed by granular carbon adsorption.

Estimates of the concentrations of BOD, COD, TOC, nitrogen, and phosphorus downstream of each group of processes are shown in Table 3. Estimated capital and operating and maintenance costs for each process are shown in Figures 2 through 7. The cost for any group of processes can be found by adding the cost for the individual processes in the group.

Table 3

ESTIMATED WATER CONTAMINANT CONCENTRATIONS IN

EFFLUENT STREAM FROM VARIOUS GROUPS OF TERTIARY PROCESSES

	VSS mg/1	BOD_5 mg/1	COD mg/1	TOC mg/1	Nitrogen* mg/1	Phosphorus mg/1 as P	Remarks
0. Secondary Effluent	20	13	60	20	17	10	
1. Microscreening or Rapid Sand Filtration (1a, 2, 7a)	6	7.5	47	16	17	10	70% Removal of Suspended Solids
2. Granular Carbon Adsorption (1c, 4, 5, 7c)	2	2	10	3	17	10	90% Removal of Suspended Solids
3. Lime Clarification (1b, 2, 3, 3a, 4, 7b)	2	6	44	15	17	1	90% Removal of Suspended Solids
4. Lime Clarification + Multi-Media Filtration (1b, 2, 3, 3a, 6, 5, 7c)	<1	5	42	14	17	1	99% Removal of Suspended Solids
5. Lime Clarification + Ammonia Stripping (1b, 2, 3, 4, 7b)	2	6	44	15	2	1	90% Removal of Suspended Solids
6. Lime Clarification + Ammonia Stripping + Granular Carbon Adsorption (1b, 2, 3, 4, 5, 7c)	<1	1	9	3	2	1	99% Removal of Suspended Solids

*Dissolved

FILTRATION AND MICROSCREENING

As a tertiary process, filtration has application first as a roughing filter which is competitive with the microscreening process and as a polishing filter which would normally be used downstream of the lime clarification process.

The roughing filter has been investigated by Truesdale and Birkbeck[1] in England and by the Metropolitan Sanitary District of Greater Chicago[2]. The roughing filter used by Truesdale and Birkbeck removed about 60% of the suspended solids from a secondary effluent containing 17 mg/l of suspended solids. Microscreening equipment investigated by Truesdale and Birkbeck in the same study removed 60% of the suspended solids also. The backwash water used was about 5% of the throughput for both the filter and the microscreen. Backwash water is returned to the secondary process.

Lynam, Ettelt, and McAloon[2] at the Metropolitan Sanitary District of Greater Chicago made measurements on a microscreening unit and a roughing sand filter. The suspended solids in the secondary effluent averaged about 11 mg/l. The average removal for the microscreen process was 70% as compared to 75% for the sand filter. It was found at Chicago that the principal filtering effect was obtained by the cake of suspended solids held on the microscreen. The speed of the drum must, therefore, be reduced as the suspended solid concentration in the influent stream is reduced. Capital cost estimates for both microstrainers and filters were made by the Chicago engineers and were found to be roughly equivalent.

The cost estimates shown in Figure 2 for microscreening were derived from a mathematical model developed at the Federal Water Pollution Control Administration's (FWPCA Taft Water Research Center (TWRC)) at Cincinnati based on the work of Boucher[3].

Microscreening of secondary effluent was also investigated by the Department of Water and Power of Los Angeles at the Hyperion Treatment Plant. With an average suspended solids concentration in the secondary effluent of 21 mg/l, it was observed that about 65% of the suspended solids was removed by the microscreen.

In tests on a microscreening unit at Lebanon, Bodien and Stenburg[4] reported suspended solids removals of 89% with a fine mesh screen and 73% with a coarser screen. Influent suspended solids averaged 17 mg/l with the fine screen and 27 mg/l with the coarser screen. BOD reduction averaged 61% for the coarse screen and 81% for the fine screen.

To summarize, the microscreening and roughing filters are about equal in both performance and cost.

Multimedia polishing filters which will remove essentially all of the suspended solids from water can also be used downstream of activated sludge or lime clarification. These filters can be used with or without the addition of chemicals such as alum or poly-electrolytes. Examples of this type of filter are the Microfloc process* used at South Tahoe Public Utility District, the Zurn

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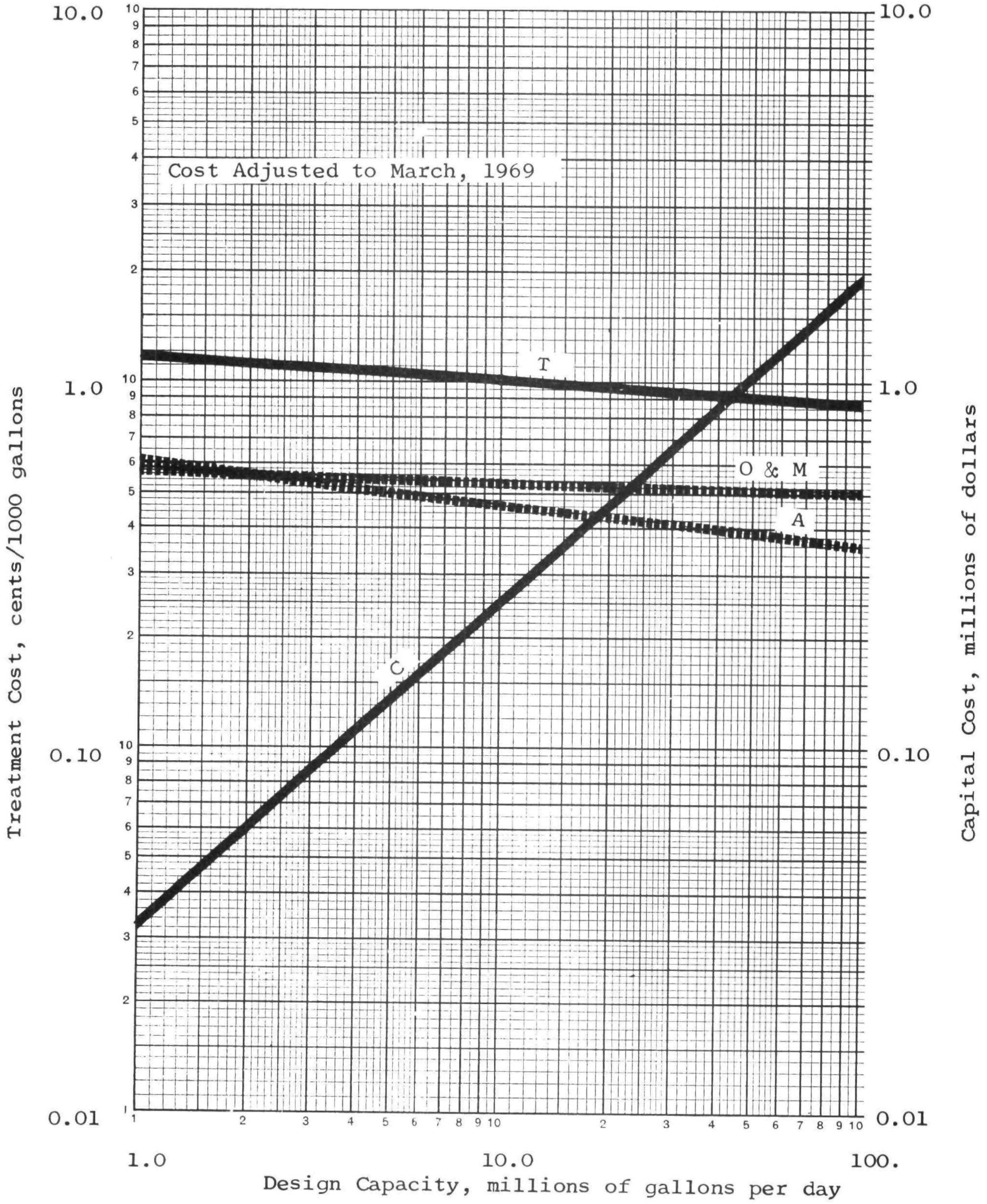
*Mention of products and manufacturers is for identification only and does not imply endorsement by the Federal Water Pollution Control Administration and the U. S. Department of the Interior.

MICROSCREENING OF SECONDARY EFFLUENT

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



C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons (4½% - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 2

filter which was tested at the TWRC pilot plant in Lebanon, Ohio, and the multimedia filters used both at Lebanon and the TWRC pilot plant at the Blue Plains Plant in Washington, D. C. No chemicals are added when the filter is used downstream of the lime clarification process. The purpose of the filter in this application is to remove inorganic fines which can cause considerable turbidity which, in turn, would be undesirable for any reuse applications. At Tahoe about 200 mg/l of alum was added upstream of the filter. Tests made on the Zurn multimedia pressure filter at Lebanon demonstrated good performance with the addition of 12.5 mg/l of alum and 2.5 mg/l of C-7 polyelectrolyte.

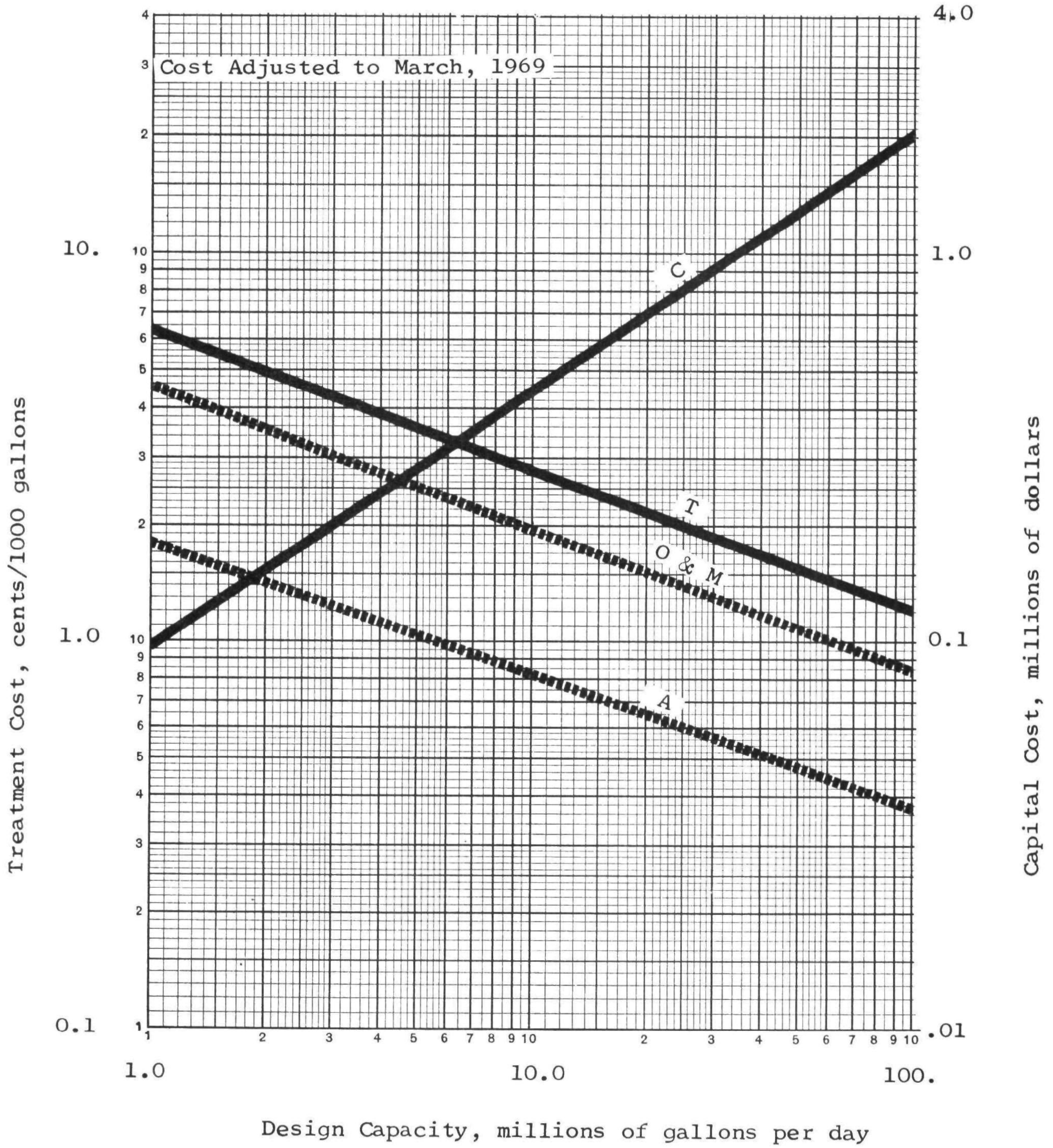
To summarize, the multimedia polishing filter is necessary for the removal of turbidity when high quality water is required. In treating water for discharge to a natural stream, the use of this filter is probably not justified. Cost estimates (from reference 5) for the multimedia filter without addition of chemicals are shown in Figure 3. Operating and maintenance costs has been reduced by 1/3 to account for the fact that settling basins are not used.

LIME CLARIFICATION

The lime clarification process is used primarily for removal of phosphorus and suspended organic matter. An additional benefit is the increased pH resulting from lime addition which makes ammonia nitrogen available for removal by air stripping. The type of equipment used is an upflow clarifier with recirculation of lime sludge. For hard water applications, one upflow clarifier appears

FILTRATION THROUGH SAND OR GRADED MEDIA - 4GPM/SQ FT

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



C = Capital Cost, millions of dollars
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O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 3

to be sufficient. For soft water, the most promosing arrangement appears to be two upflow clarifiers in series with the ammonia stripping tower downstream of the first clarifier, followed by a recarbonation unit. It is expected that the carbon dioxide required for recarbonation can be salvaged from the recalcination of the waste lime sludge.

Lime requirements to reach any target pH can be reliably calculated if the ionic character of the wastewater is known. In general, waters with high hardness require less lime and operate with lower pH values. Also some coagulant salt, such as ferrous sulfate, may be needed to help settle the inorganic fines which would otherwise escape in the effluent stream.

A 75 gpm unit has been operating at the Lebanon pilot plant for more than one year. Calcium and magnesium concentrations in the feed stream averaged 105 mg/l and 29 mg/l respectively. The TOC of the feed stream averaged 28 mg/l. The concentration of phosphate entering the process averaged 30 mg/l. Phosphate in the effluent averaged about 2.2 mg/l. The removal of BOD averaged 86%, while the removal of TOC and COD averaged 58%. An additional 5% of TOC was removed by the dual media (anthrafilt and sand) downstream of the clarifier. Good results were obtained by raising the pH to about 9.0 by the addition of about 250 mg/l of hydrated lime.

A second lime clarification pilot unit has been operated at the TWRC Blue Plains pilot plant at Washington, D. C. The feed stream there contains 40-50 mg/l of calcium and 7-10 mg/l of magnesium, which is characteristic of soft water. The organic content of the feed stream is also high, since the activated sludge process upstream is of the modified type. The BOD averages about 45 mg/l. Two upflow clarifiers in series with recarbonation have been found likely to give the best results. Ferrous sulfate can be used as a coagulant aid in the second clarifier. This arrangement reduces the BOD level to about 15 mg/l. Similarly, a 50-60% reduction in TOC has been achieved, resulting in a level of 14-15 mg/l in the effluent stream. The lime requirement to raise the pH to 11.5-12 is 350 mg/l as CaO. A dual media filter downstream of the twin clarifier unit reduces the TOC by an additional 2 mg/l. Phosphate concentrations are reduced by 93%, giving an effluent concentration of about 1.5 mg/l.

The installed cost of equipment was taken from cost estimates for Infilco Densators supplied by Infilco/General American Transportation Corporation, Tucson, Arizona. The Densators were sized for an overflow rate of 1000 gpd/sq. ft. at the mean flow rate. At least one extra Densator was provided for shut-down to provide for cleaning and repair. For example, at 1 mgd mean flow, two 40 ft. dia. Densators were sufficient, but the cost is based on three. For the 10 mgd plant size (mean flow), two 120 ft. dia. Densators were sufficient, but three were provided. At the 100 mgd size, twelve 145 ft. dia. Densators were required, but 15 were provided. The cost of lime feeding equipment was then added to obtain the total equipment cost. It was visualized that the ammonia stripping tower would be built over the second clarifier and that recarbonation for pH adjustment would be accomplished in the second clarifier. The cost of recarbonation equipment was considered minor and was not included in the equipment cost. Twenty percent of the equipment cost was added to provide for engineering and contingencies giving the total capital cost shown in Table 4 and Figure 4.

Debt service was computed as 6.744% of the total capital cost per year corresponding to interest at 4½% over a 25-year period. For operating labor, it was estimated that 12 man-hr/day/mgd would be sufficient at the 1 mgd size and that 0.2 man-hr/day/mgd would be adequate at the 400 mgd size. The estimate for operating labor at the 400 mgd size was supplied by Infilco. These two points were used to find the following relationship for operating labor.

$$\text{Operating Labor, man-hr/day/mgd} = 12(\text{mgd})^{-.68}$$

For maintenance labor, 3 man-hr/day/mgd was believed to be sufficient. Electrical power requirements were obtained from Infilco and converted to cost by taking the cost of power as 1 cent/kw-hr.

There is some evidence that a coagulant aid such as iron might be required in the second clarifier. Since the need for this chemical is not clearly established, it was not included in the cost. 5 mg/l or iron, however, can be provided for about 0.44 cents/1000 gallons.

Table 4

COST OF THE TWO CLARIFIER LIME CLARIFICATION PROCESS FOR SECONDARY EFFLUENT

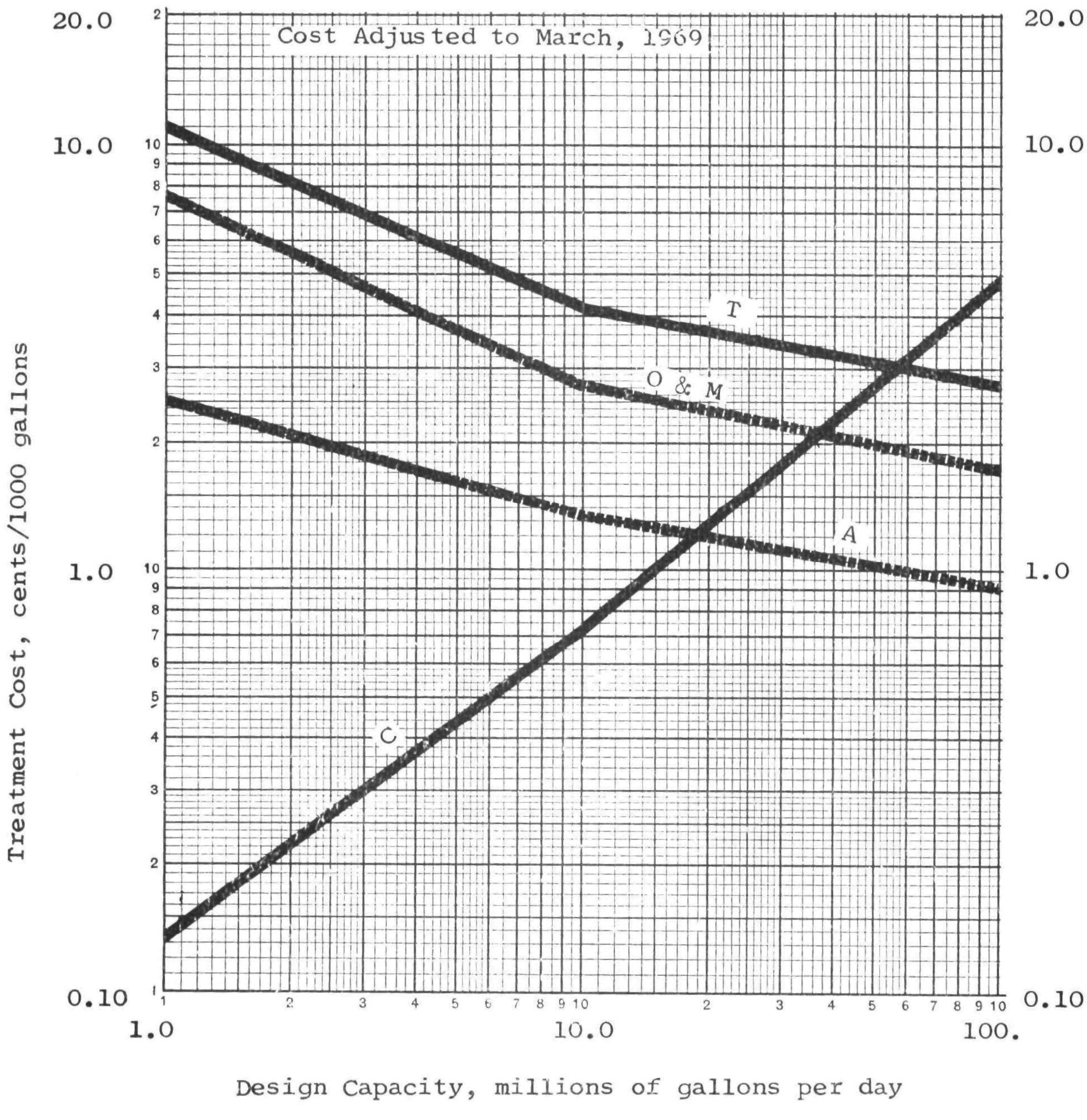
	1 mgd	10 mgd	100 mgd	309 mgd
TOTAL CAPITAL COST, dollars	138,900	721,200	4,922,000	12,200,000
PROCESS COSTS, cents/1000 gallons				
1. Amortization, 4½% and 25 yr.	2.57	1.33	.909	.730
2. Operating Labor	4.57	.952	.198	.092
3. Maintenance Labor	.942	.942	.942	.942
4. Supervision and Payroll Overhead*	1.65	.558	.339	.310
5. Maintenance Materials**	.314	.314	.314	.314
6. Electrical Power	.05	.05	.05	.05
TOTAL COST WITHOUT CHEMICALS	10.10	4.15	2.75	2.43
LIME RECALCINATION AND MAKE UP	9.17	3.67	1.92	1.40
TOTAL TREATMENT COST WITH RECALCINATION	19.27	7.82	4.67	3.83
1. Cost of Lime Delivered, 350 mg/l	2.70	2.70	2.70	2.70
2. Cost of Sludge Disposal (hauling to land fill, 25 mile one-way trip)	.67	.67	.67	.67
TOTAL TREATMENT COST WITH DISPOSAL OF LIME SLUDGE TO LAND FILL	13.47	7.52	6.12	5.80

* Taken as 30% of operating and maintenance labor

** Taken as 1/3 of maintenance labor (maintenance cost = 75% labor + 25% materials)

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 (4½ - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 4

The total installed cost of lime recalcination facilities including thickener, centrifuge, and furnace or kiln was obtained from the following existing and planned plants:

Lake Tahoe	$551,571	7.5 mgd
Piscataway, Md.	$495,600	5.0 mgd
Dayton, Ohio	$2,500,000	125 mgd
Lansing, Mich.	$1,500,000	62.5 mgd

Plotting these values for total capital cost, the following relationship is found:

$$\text{Total Capital Cost, dollars} = \$200{,}000\ (\text{mgd})^{.50}$$

The data from water treating plants such as Dayton, Ohio, Miami, Florida, and Lansing, Michigan were converted to equivalent wastewater treating installations by assuming that the ratio of lime produced in tons/day to mean flow in mgd was (1.2). This ratio was derived from experience at Lake Tahoe where 9 tons/day of recalcined lime is produced in a 7.5 mgd wastewater treating plant. Actual data on operating manpower obtained from Dayton, Lansing, Mich. and Miami, Florida were fitted with the following relationship:

$$\text{Operating Labor, man-hr/day/mgd} = 7.8\ (\text{mgd})^{-.63}$$

The maintenance labor reported was found to average about 0.6 man-hr/day/mgd. Electrical power was found to be roughly proportional to size. The cost of fuel, however, showed a marked reduction with size, which might be attributed to the savings in heat loss or the reduced price of fuel at the larger size plants. Make up lime was computed using the assumption that 350 mg/l of lime (CaO) is required and that

1.2 tons/day/mgd is recovered through recalcination. The cost of purchasing lime was taken as $18.50 per ton. Cost estimates shown in Table 5 and Figure 5 represent the complete cost of supplying lime to the lime clarification process. Total cost for the lime clarification process can be found by adding the values shown in Figures 3 and 4 as demonstrated in Table 4.

AMMONIA STRIPPING

If an ammonia stripping tower is used with the lime clarification process, the ammonia nitrogen can be removed at moderate cost. Even though many problems associated with the use of this process remain to be solved, it is presently viewed as the most promising process for removing ammonia nitrogen from wastewater. For example, the performance is strongly dependent on air and water temperature so that use of the process may not be feasible during the winter months when the temperature of ambient air is below 32°F. The probable destruction of lignin in the wooden packing of the ammonia stripping tower as a result of prolonged contact with high pH water possibly may be corrected by substituting plastic or plastic-covered wood for the normally used wooden packing. During the summer months, when nitrogen removal is usually most important, the efficiency of ammonia stripping should average about 90% removal of ammonia nitrogen.

A pilot ammonia stripping tower has been operated at Lake Tahoe under summer conditions with greater than 90% removal of ammonia nitrogen. A larger ammonia stripping tower has been recently installed (3.5 mgd) at Lake Tahoe but performance much below 90%

Table 5

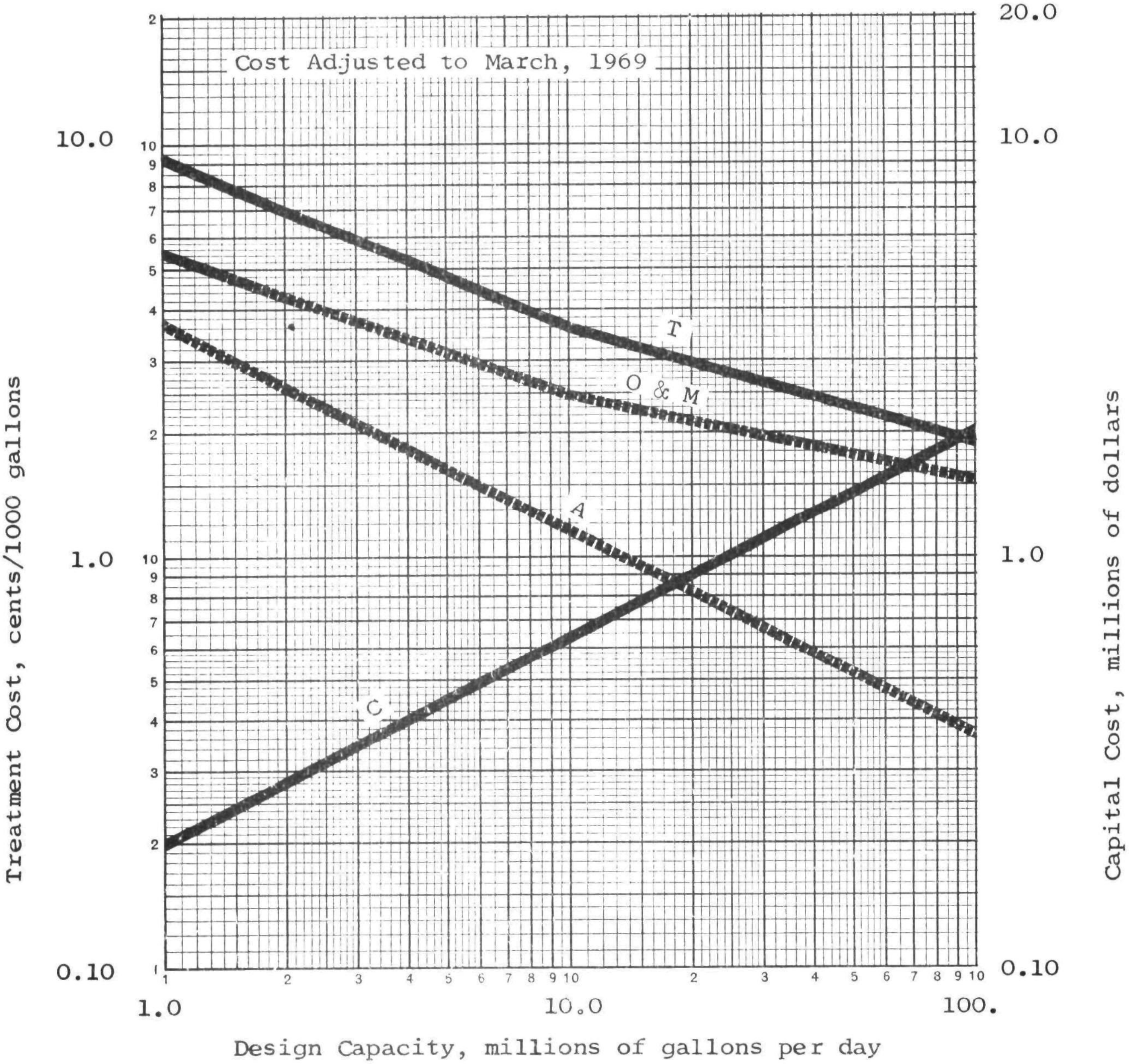
RECALCINATION OF LIME SLUDGE FROM LIME CLARIFICATION PROCESS

	1 mgd	10 mgd	100 mgd	309 mgd
TOTAL CAPITAL COST, dollars	200,000	640,000	2,000,000	3,550,000
PROCESS COST, cents/1000 gallons				
1. Amortization, 4½% and 25 yr.	3.70	1.18	.37	.066
2. Operating Labor	2.5	.595	.138	.066
3. Maintenance Labor	.190	.190	.190	.190
4. Supervision and Payroll Overhead*	.806	.236	.110	.077
5. Maintenance Materials**	.063	.063	.063	.063
6. Electrical Power	.100	.100	.100	.100
7. Fuel	1.33	.824	.47	.353
8. Make Up Lime, $18.50/ton	.48	.48	.48	.48
TOTAL RECALCINATION COST, cents/1000 gallons	9.17	3.67	1.92	1.40
RECALCINATION PLANT CAPACITY, tons/day	1.2	12.0	120.0	371.
RECALCINATION COST, dollars/ton	72.4	26.5	12.0	7.6

* Taken as 30% of operating and maintenance labor
** Taken as 1/3 of maintenance labor (maintenance cost = 75% labor + 25% materials)

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 (4½ - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

Figure 5

removal has been experienced under winter operating conditions. Pilot ammonia stripping towers are also being installed at the Hanover Plant by The Metropolitan Sanitary District of Greater Chicago but no tests results are, as yet, available.

A computerized design procedure for estimating performance and cost of ammonia stripping towers has been completed by the Illinois Institute of Technology Research Institute under contract to FWPCA.

Only one value for installed cost was found for ammonia stripping towers. At Lake Tahoe, the installed cost of the tower was $224,500 and the cost of the concrete basin was $100,500 giving a total cost of $325,000 for an installation sized at 3.75 mgd. Since it was learned in talks with the Marley Co. that little economy of size is realized for cooling towers, a capital cost line through this point was drawn with a slope of (0.90). Bechtel Corp. estimated the cost of a 30 mgd ammonia stripping installation as $2,575,000. This point fell only slightly above the line with a 9/10 slope. Amortization was taken as 4½% over a 25 year period, but this is assuming that the film packing will not have to be replaced within the assumed period.

In talks with engineers at Lake Tahoe the opinion was expressed that the packing might have to be replaced in ten years. Marley Co. estimated that if the packing is replaced, the cost will be between 50% and 60% of the installed cost. Marley Co., however, feels that if additional large ammonia stripping towers are built, the life can be extended by using improved materials for packing. Since this problem is totally unresolved, no attempt was made to account for the additional cost.

Operating labor was estimated by assuming that 8 man-hr/day would be sufficient at the 1 mgd size and that 40 man-hr/day would suffice at the 100 mgd size. A line through these two points is represented by the following relationship:

$$\text{Operating Labor, man-hr/day/mgd} = 8.0\ (\text{mgd})^{-.65}$$

Maintenance labor was estimated at 1.5 man/hr/day/mgd. Electrical power was found by scaling up the Lake Tahoe plant which has a 100 horsepower fan and a 30 horsepower water pump. Cost of power was taken as one cent/kw-hr. All costs associated with the process are shown in Table 6 and Figure 6.

GRANULAR ACTIVATED CARBON

Most of the practical operating experience with the granular carbon adsorption process for treating secondary effluent has been gained at the Pomona, California pilot plant operated jointly by FWPCA and the County Sanitation Districts of Los Angeles County. This pilot plant, which has a design flow of 288,000 gpd, consists of five downflow pressure contactors, four of which are normally in operation. The contact time is 36-40 minutes. No pretreatment of secondary effluent from the activated sludge plant is used. Secondary effluent, however, is of high quality; 10 mg/l suspended solids, 47 mg/l COD, and 13 mg/l TOC. Backwash water (8000 gallons) is used once a day to bachwash the first contactor. This represents about

Table 6

COST ESTIMATES FOR AMMONIA STRIPPING OF LIME CLARIFIED WASTEWATER

	1 mgd	10 mgd	100 mgd	309 mgd
TOTAL CAPITAL COST, dollars	95,000	760,000	6,000,000	17,000,000
PROCESS COSTS, cents/1000 gallons				
1. Amortization, 4½% and 25 yr.	1.76	1.40	1.11	1.02
2. Operating Labor	2.51	.55	.126	.060
3. Maintenance Labor	.471	.471	.471	.471
4. Supervision and Payroll Overhead*	.894	.306	.179	.159
5. Maintenance Materials**	.205	.205	.205	.205
6. Electrical Power	.69	.69	.69	.69
TOTAL TREATMENT COST, cents/1000 gallons	6.53	3.62	2.78	2.61

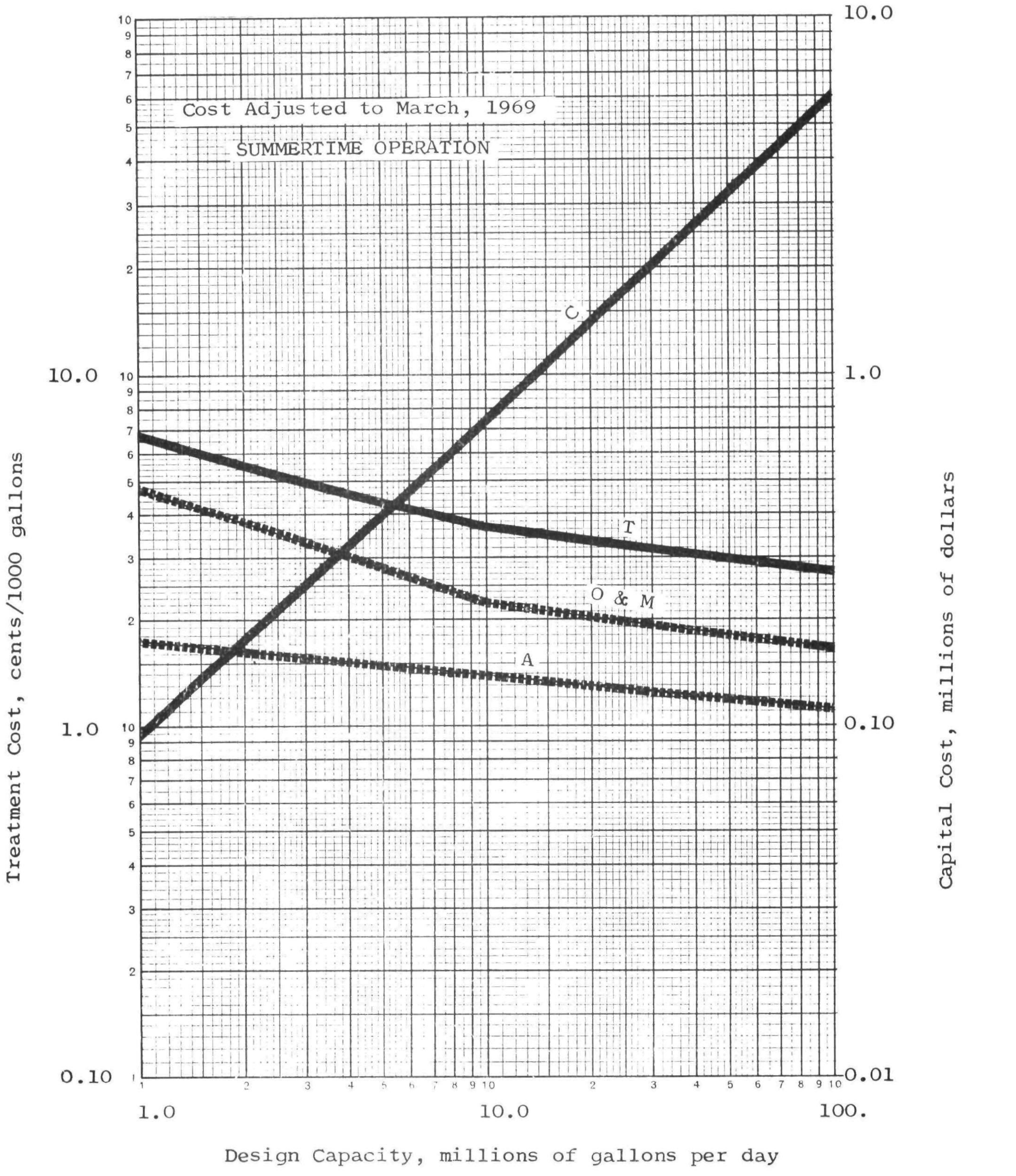
Note: Wages for Water, Steam, and Sanitary Systems Nonsupervisory Workers for March, 1969 = $3.14/hr.

* Taken as 30% of operating and maintenance labor

** Taken as 1/3 of maintenance labor (maintenance cost = 75% labor + 25% materials)

AMMONIA STRIPPING PROCESS

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



C = Capital Cost, millions of dollars
A = Debt Service, cents per 1000 gallons (4½ - 25 yr.)
O & M = Operating and Maintenance Cost, cents per 1000 gallons
T = Total Treatment Cost, cents per 1000 gallons

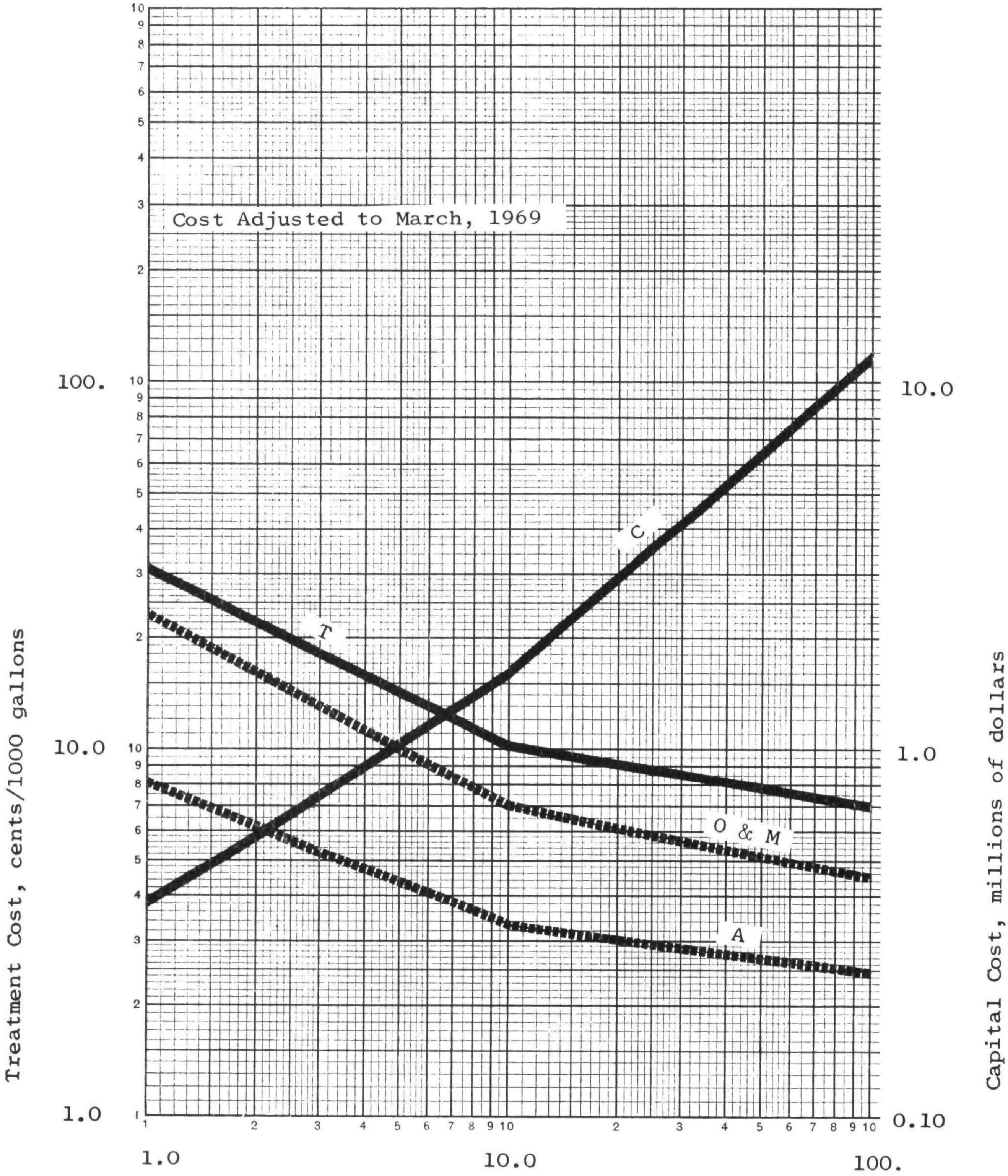
Figure 6

3% of the throughput. The wash water is returned to the secondary plant. It is estimated that if 20 mg/l of suspended solids are applied, the backwash will be required twice a day using about 6% of the throughput. The suspended solids concentration in the effluent stream is normally less than 1 mg/l. About 80% of the organic species (COD, TOC) are normally removed.

A recent preliminary design study by the M. W. Kellogg Company under contract to FWPCA estimated that a design based on 50 minutes contact time, would reduce a COD of 60 mg/l in the feed stream to 7 mg/l in the effluent stream. Estimated removal values shown in Table 3 are based partially on this study. Cost estimates shown in Figure 7 are also based on calculations of the Kellogg Company which provides downflow, pressure contactors with 50 minutes contact time.

GRANULAR CARBON ADSORPTION PROCESS

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



C = Capital Cost, millions of dollars

A = Debt Service, cents per 1000 gallons (4½% - 25 yr.)

O & M = Operating and Maintenance Cost, cents per 1000 gallons

T = Total Treatment Cost, cents per 1000 gallons

Figure 7

REFERENCES

1. Truesdale, G. A. and Birkbeck, A. E., "Tertiary Treatment of Activated Sludge Effluent", International Filtration Conference, Water Pollution Research Laboratory, Stevenage, Herts (1967).

2. Lynam, B., Ettelt, G. and McAloon, T., "Tertiary Treatment at Metro Chicago by Means of Rapid Sand Filtration and Micro-Strainers" Presented at WPCF 41st Annual Conference, September, 1968.

3. Boucher, P. L., "A New Measure of the Filterability of Fluids with Application to Water Engineering", ICE Journal 24, pp. 415-446, (1947).

4. Bodien, D. G. and Stenburg, R. L., "Microscreening Effectively Polishes Activated Sludge Plant Effluent", Water and Wastes Engineering, Vol. 3, No. 9, September, 1966.

5. Smith, Robert, "Cost of Conventional and Advanced Treatment of Wastewaters", Jour. WPCF (Annual Conference Issue) Vol. 40, No. 9, pp. 1546-1574, September, 1968.