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DIGITAL COMPUTER SIMULATION OF WASTEWATER TREATMENT

Paul W. Cummings, Jr. and
Donald Dean Adrian, Project Investigator

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by

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Dr. John H. Nebiker, formerly Assistant Professor of Civil Engineering, now Vice President, Curran Associates, Inc., Northampton, Massachusetts, was Mr. Cummings general adviser in the masters program.

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ABSTRACT

Computer simulations of modifications to the standard activated sludge process are developed and integrated into a model developed by the Federal Water Pollution Control Administration. The modifications developed are: deletion of primary treatment; aerobic sludge digestion; and a more complete routine for sand bed sludge dewatering. Cost estimates associated with the modifications are presented and compared with the standard system of activated sludge.

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PART I: INTRODUCTION.

1.1 Need for the Study

The ever expanding population of this country, and the shift of the population from rural areas to the large urban centers has brought about an urgent need for the protection of the nation's water resources. The federal government has taken the initiative, and through the Department of the Interior has instituted a large scale attack upon water pollution.

This program for cleaning the nations waterways must of necessity include a large program for the construction of public works designed to treat the always present sewage flow of the population. There are available to the design engineer several different treatment processes, and combinations of processes. Grouped into two broad categories, there are: primary sewage treatment; and secondary sewage treatment. There are, of course, also tertiary treatment, and chemical treatment processes, but these do not enjoy any large, widespread role in the present system of sewage treatment works. The present direction now seems to be toward universal secondary treatment.

The design engineer must be able to select the type of treatment best suited to the situation, and the regulatory agencies have a need to supervise the work of the design engineer, and also the needs of the whole area or watershed. The regulatory agencies are also often involved in the funding of construction projects, and in the financing of the projects. For these reasons, both the design engineer, and the regulatory and planning agencies require a means of selecting proper treatment systems.

One alternative in the decision process, the computer simulation, will be the topic of the following investigation.

1.2 Purpose and Objectives of the Study

The purpose of this study is to obtain a working model of the activated sludge sewage treatment process, both in its standard form, and with options for various alternative processes and process arrangements.

The objectives in the study are to achieve the capability of predicting the treatment costs for the various configurations of the activated sludge process. This capability would give planners, designers and local officials a rapid efficient means of selecting a total treatment scheme intelligently, by looking at as many alternatives as possible without the great expense of complete preliminary designs for each alternative.

1.3 Scope of the Study

This study will attempt to modify an existing model (21), for the standard activated sludge process to include treatment schemes which will: delete primary sedimentation from the standard activated sludge process; replace anaerobic sludge digestion with aerobic sludge digestion; and develop an optimizing routine for the design of sand drying beds for sludge dewatering. Total treatment costs will be the basis for the comparisons, with the standard for comparison being the standard activated sludge process costs. The effluents of each type of treatment scheme will meet the same BOD and Suspended Solids standards.

Finally, a comparison will be made between the computer simulation and an actual treatment plant located in Lee, Massachusetts. The Lee treatment plant does not employ primary sedimentation, utilizes aerobic sludge digestion, and dewaters the digested sludge on sand drying beds.

PART 2: BASIC CONSIDERATIONS

2.1 Deletion of Primary Sedimentation

Almost all secondary sewage treatment plants are designed to be an addition to the basic primary sedimentation process. The standard activated sludge units are connected to the primary sedimentation tanks, and sometimes the secondary sludge is routed through the primary tank to increase the overall sludge solids concentration. Trickling filters are also typically preceded by the primary sedimentation process, which removes the gross settleable solids.

The exception is found in small extended aeration plants, and other modifications of the activated sludge process as found in many designs of small "package" plants. It is thought in these small plants that simplicity and economies of construction make it worth while to omit the primary sedimentation phase, especially in light of the need to minimize labor and maintenance.

Contact stabilization, as used in small plants, aerates the sludge from the secondary settler, and returns it to the aerator to be mixed with the main flow. Larger contact stabilization plants often do utilize primary sedimentation.

Extended aeration plants utilize long detention periods in the aerator to completely oxidize the organic material in the raw sewage. Detention times range from ten to twenty-four hours.

One of the aims of this research is to investigate the effects of deleting primary sedimentation from medium and large sized treatment plants.

One of the first and most obvious effects to be expected is an increase in the required detention time in the aerator so as to oxidize the sewage to a degree equivilant to the results of a system in which primary sedimentation is utilized. This will affect the size of the aerator, the capacity of the air blowers, and the ratio of the mixed liquor active suspended solids to the total suspended solids in the aerator.

Another topic of importance would be the effect of deleting primary sedimentation upon the quality of the resultant sludge. The sludge from a standard activated sludge system is a mixture of highly volatile, relatively untreated solids, and of a thin, biologically active sludge mass that has been produced by aerobic assimilation of organic material by a mixed microbiological population. The sludge from an extended aeration plant would not contain the highly volatile untreated solids. This difference will be of significant importance in the case where aerobic digestion is employed.

Assuming that the performance of the two types of plants will be comparable, the deciding factor will be economic. This will be the major consideration here, to determine if it is possible to economically treat the sewage without primary sedimentation, for it is reasonable to say that the more simple the plant is, the easier it is to operate efficiently. If the goals can be met by two different plants economically, the simpler design is the better design.

2.2 Aerobic Digestion

Aerobic digestion is the oxidation of volatile, biodegradable organic material by microbiological organisms which utilize free dissolved oxygen. This is in contrast to the more standard anaerobic digestion, a process which is carried on by microbiological populations which do not require free dissolved oxygen.

Anaerobic digestion is now the standard and almost universal digestion method employed. It is optimally operated in the mesophilic temperature range, at about 98°F., and is completely mixed if it is of the high-rate type. Detention periods for high-rate digestion are generally from fifteen to twenty-five days for the first stage where the actual digestion takes place. Usually a second unheated, unmixed digester is provided for settling.

Methane and carbon dioxide are the principal products, the methane often being utilized to heat the digester. The bacterial population responsible for the oxidation fall into two large groups: the organic acid formers, and the methane producers. The system in which these two groups exist is quite unstable, and the methane formers are quite sensitive to the pH of the system, which in turn is controlled by the activity of the organic acid forming group. The methane forming bacteria require the organic acids produced by the acid formers, but can only utilize the acids under certain environmental conditions.

Therefore, it can be seen that anaerobic digestion can be a complicated system to control, and indeed, digester upset is not uncommon.

Aerobic digestion offers the possibility of a more stable process, utilizing a bacterial population already cultivated by the activated sludge process. More importantly, it operates on types of equipment which are fairly simple, and already present in the activated sludge aeration tanks. It eliminates the necessity for gas collection, high temperatures, and close process control.

It will be part of the investigation to simulate the aerobic digestion process mathematically on the digital computer, and attempt to analyze the economic feasibility of the process.

2.3 Sludge Dewatering on Sand Drying Beds

Once the sewage solids have been separated from the main flow in a primary plant, and once the soluble pollutants have been synthesized into cellular material by one of the secondary biological processes and removed from the main stream, the solids must still be disposed. The solids have been concentrated by the treatment of the sewage, by a factor ranging from twenty to one-hundred or more. However, the solids are still water-borne, and must be dewatered before ultimate disposal.

One of the oldest methods for sludge dewatering has been by placing the sludge on sand beds to allow the water to both filter through the sand, leaving the solids behind, and to evaporate, also leaving the solids behind. Two other, and newer, methods for sludge dewatering are vacuum filtration, and centrifugation. Both of these methods require the input of power, and almost always the addition of chemical conditioners to promote a good cake formation. Both of these methods also have large capital costs associated with them.

Sand drying beds also have disadvantages associated with them, such as decreased performance during cold weather; slowness of the drying; large land requirements; and relatively little effort to improve the process, for there are no salesmen for sand drying beds.

This investigation will attempt to formulate a model of the sand bed dewatering process which will be fit into the general scheme of the activated sludge treatment simulation. An attempt to optimize the design and operation of the drying beds will be made, and a comparison with the cost of vacuum filtration will also be made.

PART 3: LITERATURE REVIEW

3.1 Previous Simulation Techniques

The Federal Water Pollution Control Administration (FWPCA) has developed a simulation of the basic activated sludge process, under the direction of Robert Smith (21) at the Cincinnati Water Research Laboratory. The present investigation will use that model as a basis. Modifications and additional processes and configurations will be added to the FWPCA model.

Since the FWPCA model, titled: Preliminary Design and Simulation of Conventional Wastewater Renovation Systems Using the Digital Computer, will be the basis for this new work, a fairly complete description of the model will be presented here.

The basic process configuration is shown in Figure 1, with the exception that there is a provision for sand drying beds contained in the program. The only two decisions concerning treatment process selection are whether vacuum filters will be used instead of sand drying beds, and whether the supernatant stream from the solids handling processes is to be returned to the head of the plant.

3.1.1 Waste Characteristics

The influent waste is characterized in two main groups, solid and dissolved species. Each of these groups is divided into: degradable carbon; non-degradable carbon,

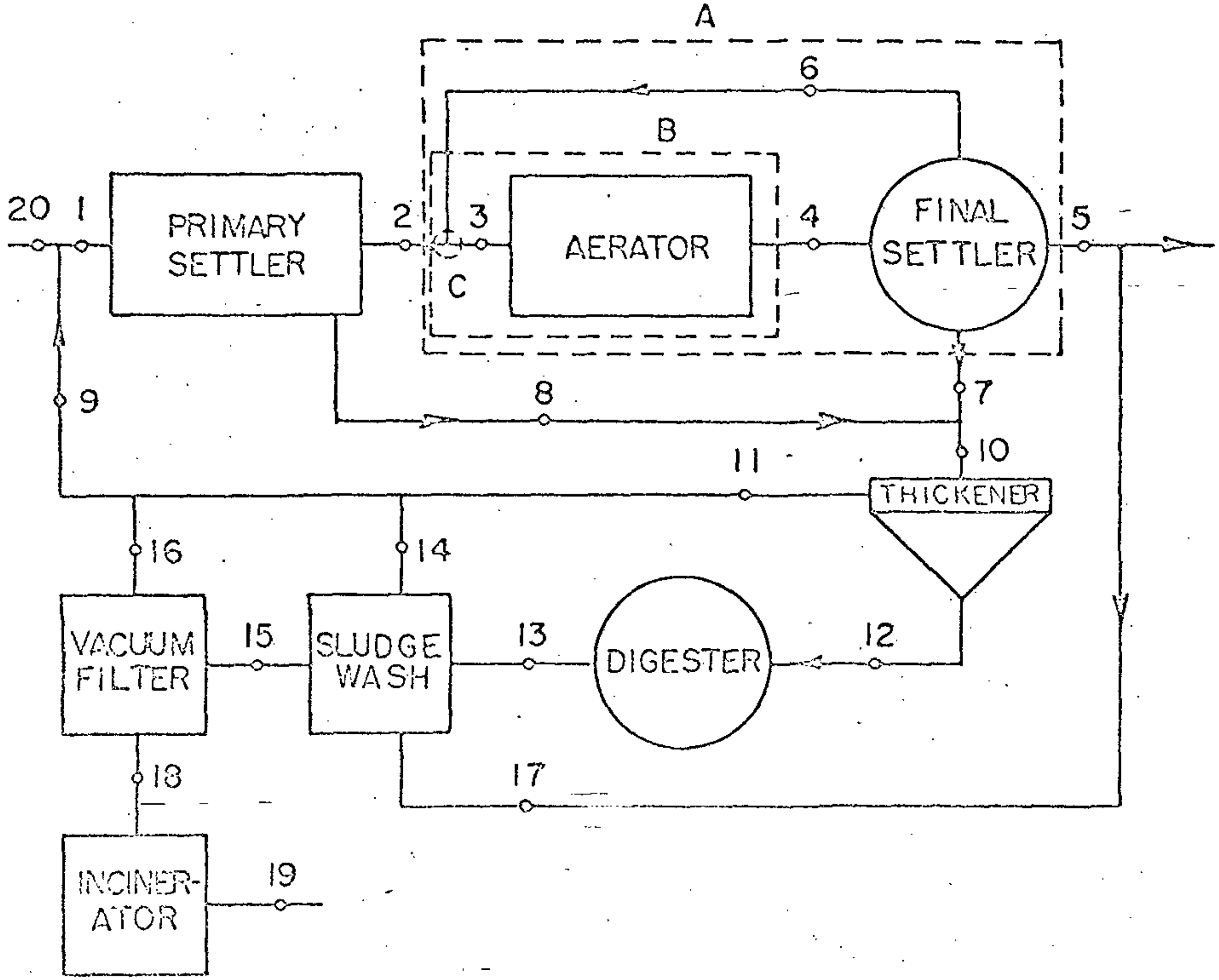

Figure 1.

CONVENTIONAL PROCESSES SYSTEM DIAGRAM

nitrogen; phosphorous; and inorganic fixed matter.

The following ratios are used to relate several test results to each other:

	raw sewage	activated sludge
BOD/TOC =	1.87	-
COD/VSS =	1.50	1.42
COD/TOC =	3.2	2.7
BOD/VSS =	-	0.84

where:

BOD = Biochemical Oxygen Demand

TOC = Total Organic Carbon

COD = Chemical Oxygen Demand

VSS = Volatile Suspended Solids

TSS = Total Suspended Solids

3.1.2 Primary Settler

The model assumes that only solid material is removed in the primary settler, and that all of the different classes of solids are removed in the same manner. Input to the program are the fraction of solids to be removed by the settler, and the degree of thickening obtained in the settler. From this information, the two effluent streams from the settler are characterized: the settled sewage, and the sludge underflow.

The overflow rate for the primary settler is determined by a curve fit through data obtained from full sized plants:

$$FRPS = 0.82\ e^{-GPS/2780} \qquad (1)$$

where: FRPS = fraction of the solids removed

GPS = overflow rate gal/day/sq ft

3.1.3 Aerator and Final Settler

The performance of the aerator and final settler are determined together as related processes. The size of the aerator is found by the following equation assuming that all the BOD leaving the aerator is in the dissolved form:

$$VAER = \frac{Q_2\ (BOD_2 - BOD_5)}{CAER \times MLSS \times BOD_5} \qquad (2)$$

where: VAER = volume of aerator

CAER = rate constant

MLSS = mixed liquor suspended solids

Q_2 = flow at Station 2, M6D

An empirical formula is used to predict the performance of the final settler, developed from a bench top study at the Cincinnati Water Research Laboratory:

$$XRSS = \frac{556(GSS)^{0.494}}{MLSS^{1.82}(24 \cdot TA)^{0.439}} \qquad (3)$$

where: XRSS = fraction of solids going over weir

GSS = final settler overflow rate

TA = aerator detention time

GSS, the final settler overflow rate, is supplied as input to the simulation.

Next Smith wrote a materials balance around the aerator and the final settler, assuming that each pound of five day BOD removed in the aerator was synthesized into 0.65 pounds of active solids:

$$0.65 \cdot FOOD \cdot Q_2 - CEDR \cdot VAER \cdot MLASS = MLASS(Q_5 \cdot XRSS + Q_7 \cdot URSS) \tag{4}$$

where: FOOD = amount of BOD synthesized, mg/l/day

CEDR = fraction of active solids destroyed/day

MLASS = *mixed liquor active suspended solids*

URSS = ratio of solids content in final settler sludge to solids content in aerator

From this materials balance, it was found that the sludge volume from the final settler was:

$$Q_7 = \frac{Q_2((0.65FOOD/MLASS) - XRSS) - CEDR \cdot VAER}{URSS - XRSS} \tag{5}$$

Then, the overflow from the final settler was:

$$Q_5 = Q_2 - Q_7 \tag{6}$$

Now the increase in non-biodegradable material resulting from the destruction of the active mass in the aerator was determined:

$$0.12(FOOD)Q_2 = (MLDSS + 0.185MLASS)(XRSS \cdot Q_5 + URSS \cdot Q_7) \tag{7}$$

where: MLDSS = non-biodegradable matter in the mixed liquor suspended solids

Solving for the MLDSS, we have:

$$\text{MLDSS} = \frac{0.12(\text{FOOD})Q_2}{Q_5 \times \text{XRSS} + Q_7 \times \text{URSS}} - 0.185 \times \text{MLASS} \qquad (8)$$

An iterative technique was used to solve for the amount of FOOD synthesized each day into active solids. The MLASS is known to lie between two values: zero and a maximum of either the MLSS or the value of MLASS which would result if all the incoming BOD was synthesized into active solids. Starting with the waste sludge stream equal to zero, a bisection method of root finding was used, varying the assumed value for MLASS and FOOD. A materials balance about the aerator determined the amount of sludge to be wasted:

$$Q_6 = \frac{Q_2\left(1.0 - \frac{0.65 \times \text{FOOD}}{\text{MLASS}}\right) + \text{CEDR} \times \text{VAER}}{\text{URSS} - 1.0} \qquad (9)$$

The air requirements of the aerator were determined by the relationship:

$$\text{lb } O_2/\text{day} = (\text{COD of FOOD used}) - (\text{COD of cells produced}) \qquad (10)$$

COD of cells: $1.42(0.65\text{FOOD}\cdot Q_2 - 0.185\cdot\text{MLASS}\cdot\text{CEDR}\cdot\text{VAER})8.33$

COD of FOOD used:

$1.5(\text{FOOD}\cdot Q_2)8.33$

therefore, the amount of oxygen used per day is:

$$\text{1/b } O_2/\text{day} = 8.33(0.58\text{FOOD}\cdot Q_2 \; 1.16\text{MLASS}\cdot\text{CEDR}\cdot\text{VAER}) \qquad (11)$$

Added to the above amount is the oxygen required to achieve nitrification. According to the following chemical equation, four moles of oxygen are required to convert one mole of nitrogen to nitrate:

$$2NH_4^+ + 4O_2 \text{-----} 2NO_3^- + 2H_2O + 4H^+ \qquad (12)$$

therefore, 64./14., or 4.57 lb of oxygen are required to convert one pound of nitrogen to nitrate.

The model assumes an air density of 0.075 lb/cu ft at sea level, and the air to be composed of 23.2% oxygen. Thus the computer program finds the amount of air required daily by the formula:

$$AIRCFD \; - \; (lb/day\ of\ O_2)/0.075/0.232/AEFF \qquad (13)$$

where: AIRCFD = cubic feet of air supplied each day

AEFF = oxygen transfer efficiency

AEFF is calculated by the following equation:

$$AEFF = AEFF_{20}((DOSAT-DO)/DOSAT)\cdot(1.02)^{DEGC\ -\ 20} \qquad (14)$$

where: $AEFF_{20}$ = efficiency of oxygen transfer at 20°C, and zero dissolved oxygen

DOSAT = dissolved oxygen at saturation

DO = dissolved oxygen in the aerator

DEGC = temperature in degrees centigrade

The diffusers are assumed to be submerged fifteen feet deep, and the head loss through the piping and diffusers is assumed to be 25% of the submergance head. From this power for the blowers is found by:

$$\text{Horsepower} = \frac{SCFM \cdot \Delta P \cdot 144}{33000 \cdot EFF} \quad (15)$$

where: SCFM = standard cubic feet per minute

ΔP = pumping pressure in psi

EFF = efficiency assumed to be 0.80

3.1.4 Thickener

The sludge thickener is governed by two input parameters, TRR, the solids recovery ratio, and TSS_{12} the total suspended solids in the effluent.

$$TRR = \frac{Q_{12} \cdot TSS_{12}}{Q_{10} \cdot TSS_{10}} \quad (16)$$

thus:

$$Q_{12} = TRR \cdot Q_{10} \cdot TSS_{10}/TSS_{12} \quad (17)$$

and:

$$Q_{11} = Q_{10} - Q_{12} \quad (18)$$

The total solids in the overflow is then:

$$TSS_{11} = (1-TRR) \cdot Q_{10} \cdot TSS_{10}/Q_{11} \quad (19)$$

the various other species of solids are assumed to be present in the same ratio as the total suspended solids, and the dissolved species are assumed to remain unchanged from the inflow stream.

The size of the thickener is determined from rule-of-thumb values in terms of solids loading in pounds per day per square foot, or overflow rates in terms of gallons per day per square foot of surface area. The greater of the two values is used for the required thickener size.

3.1.5 Anaerobic Digestion

It was assumed that only biodegradable material will be changed in the digester, so the biodegradable carbon at the entrance is calculated as:

$$DIGC_{12} = SOC_{12} - SNBC_{12} + DOC_{12} - DNBC_{12} \tag{20}$$

A mass balance was written about the digester using F to represent the biodegradable carbon, and C to represent the concentration of active solids in the digester:

$$VDIG \frac{dC}{dt} = VDIG \frac{K_1 CF}{K_2 + F} - b \cdot VDIG \cdot C - Q_{13} \cdot C \tag{21}$$

where: VDIG = volume of the digester

The first term is described as the time rate of change of the active organisms. The second term is the expression for growth of active solids. The third term is the decay of active solids, and the last term is the active solids leaving in the effluent.

Since the model stipulates steady state conditions exist, the first term goes to zero, and it is said that the decay term is very small and negligible. Therefore the equation degenerates to:

$$\frac{K_1 F}{K_2 + F} = \frac{Q_{13}}{VDIG} = \frac{1}{TD} \tag{22}$$

where: TD = digester detention time

The constants, K_1, and K_2 are temperature dependent, and can be represented by:

$$K_1 = 0.28\ e^{-0.036(35-t)} \tag{23}$$

$$K_2 = 700\ e^{0.10(35-5)} \tag{24}$$

where: t = degrees centigrade, varying between 20°C. to 35°C.

Since TD, the digester detention time is specified in the input, F (the concentration of biodegradable carbon in the completely mixed digester) can be found.

The biodegradable carbon in the effluent which is dissolved is assumed to be equal to the remaining volatile acid concentration, DF. It is then assumed that the following equation can be used to find DF:

$$\frac{1}{TD} = \frac{K_1 . DF}{CDF + DF} \tag{25}$$

$$CDF = 200\ e^{0.12(35-T)} \tag{26}$$

then:

$$DOC_{13} = DNBC_{12} + DF \tag{27}$$

$$SOC_{13} = SNBC_{12} + F - DF \tag{28}$$

There is a minimum detention time below which the digestion process breaks down. This is said to be the reciprocal of K_1. A safety factor of 2.5 is provided, and if this minimum detention time is longer than the time specified in the input, the detention time is changed to equal the minimum time of $K_1/2.5$. There is also another detention time which would be the optimum time for gas production. This can vary from somewhat below the minimum time at high temperatures to above the minimum at lower temperatures. This time is predicted as:

$$TDOPT = (1-(K_2/(K_2 + DIGC_{12})^{1/2}))/K_1 \qquad (29)$$

where: TDOPT = the optimum detention time for gas production

$DIGC_{12}$ = biodegradable carbon at station 12

The minimum detention time is set as the maximum of the two values: TDOPT and $K_1/2.5$

Methane production in the digester is related to the COD reduction of the sludge as follows:

$$CH4CFD = (DIGC12 - DIGC13)(3.5)(5.61)(8.33)Q_{12} \qquad (30)$$

where: CH4CFD = daily methane production in ft^3

DIGC12 = biodegradable carbon at 12

DIGC13 = biodegradable carbon at 13

3.1.6 Sludge Elutriation

The sludge elutriation process is governed by two input parameters: the solids recovery ratio, ERR, and the solids concentration at the effluent. These values are the product of engineering judgment, and not of any scientific theory. The wash water ratio, WRE, is also input, and is the ratio of wash water to sludge, taken as 3.0 by the Smith simulation. The suggested values for ERR and TSS(15) are 0.75 and 60000, respectively. The stream values are calculated as follows:

$$ERR = \frac{Q_{15} \times TSS_{15}}{Q_{13} \times TSS_{13}} \qquad (31)$$

$$Q_{15} = ERR \cdot Q_{13} \cdot TSS_{13}/TSS_{15} \qquad (32)$$

$$Q_{17} = WRE \cdot Q_{13} \qquad (33)$$

$$Q_{14} = Q_{13} + Q_{17} = Q_{15} \qquad (34)$$

where: ERR = solids recovery ratio

WRE = Wash water ratio

The total solids concentrations are now found by:

$$TSS_{14} = Q_{13} \cdot (TSS_{13}(1-ERR) + WRE \cdot TSS_5)/Q_{14} \qquad (35)$$

The various components of the total solids are assumed to exist in the ratio present in the influent stream.

Therefore:

$$SOC_{14} = (TSS_{14}/TSS_{13})\ SOC_{13} \qquad (36)$$

$$SOC_{15} = (TSS_{15}/TSS_{13})\ SOC_{13} \qquad (37)$$

Similarly the other solids species are defined as being in the ratio of the total solids times the influent value.

The dissolved species are determined by assuming complete mixing in the elutriation tank, thus the dissolved carbon would be found by writing:

$$Q_{13} \cdot DOC_{13} \quad Q_{17} \cdot DOC_5 = Q_{14} \cdot DOC_{14} \; Q_{15} \cdot DOC_{15} \tag{38}$$

Now since $DOC_{14} = DOC_{15}$ assuming complete mixing, the dissolved carbon at either station 14 or 15 can be found as:

$$DOC_{15} = (Q_{13}/(Q_{13} + Q_{17}))DOC_{13} + (Q_{17}/(Q_{13} + Q_{17}))DOC_5 \tag{39}$$

Thus all the dissolved species are found, especially the alkalinity, which is the prime purpose of elutriation. Smith did not feel that the model for elutriation was especially good, as it relies upon engineering judgment and experience, not upon soundly developed theory or experimentatior.

3.1.7 *Vacuum Filtration*

The design parameters for the vacuum filter are the liquid loading rate in gallons/hour/sq ft, and the water content of the sludge cake. Smith uses an empirical equation to determine the final water content, based upon the original water content:

$$WP = 88.0\ (TSS_{15}/10{,}000.)^{-0.123} \qquad (40)$$

where: WP = final per cent water in the cake

Writing a mass balance about the vacuum filter, Smith found:

$$W_{s15} = W_{s16} + W_{s18} \qquad (41)$$

$$W_{w15} = W_{w16} + W_{w18} \qquad (42)$$

$$W_{w18} = WP(W_{w18} + W_{s18})/100 \qquad (43)$$

$$W_{s18} = W_{w18}\ (100 - WP)/WP \qquad (44)$$

but:

$$\frac{Q_{15} \cdot TSS_{15} - Q_{16} \cdot TSS_{16}}{1{,}000{,}000.(Q_{15} - Q_{16})} = \frac{W_{s18}}{W_{w18}} \qquad (45)$$

by combination,

$$TEMP1 = 1{,}000{,}000.\ (100 - WP)/WP \qquad (46)$$

$$Q_{16} = Q_{1t}(TEMP1 - TSS_{15})/(TEMPL - TSS_{15}) \qquad (47)$$

The total suspended solids value of the vacuum filter filtrate is input, and supplied as another engineering estimate. With the TSS_{16} known, the weight of the total solids at station 18 is found as follows:

$$W_{s18} = 8.33(Q_{15} \cdot TSS_{15} - Q_{16} \cdot TSS_{16}) \tag{48}$$

As before, the separate species of solid material are assumed to be present in the same ratio as the total suspended solids, thus:

$$SOC_{16} = SOC_{15}\ (TSS_{16}/TSS_{15}) \tag{49}$$

Again, the dissolved species are assumed to be present in the same concentration as in the influent:

$$DOC_{16} = DOC_{15} \tag{50}$$

and because the stream values at station 18 are in terms of total weight,

$$\text{lb. Dissolved carbon at 18} = DOC_{15}(Q_{15} - Q_{1y})^{8.33} \tag{51}$$

Next the area required for the vacuum filter is found:

$$AVF = \frac{Q_{16}(1{,}000{,}000.)}{24.0 \cdot VFL \cdot TVF} \tag{52}$$

where: AVF = area of vacuum filter in sq ft

VFL = loading rate in gal/hr./ft^2

TVF = fraction of each day filter is operated

The chemical demand of the sludge is based upon the alkalinity and the ratio of volatile solids to solid fixed matter:

$$Y1 = 100(ALK_{15})/TSS_{15} \qquad (53)$$

$$Y2 = VSS_{15}/SFM_{15} \qquad (54)$$

then the demand for chemical, $FECL_3$ is found by:

$$FECL3 = 1.08(Y1) + 2.0(Y2) \qquad (55)$$

$$DFECL3 = TSS_{15}(FECL3/100)Q_{15}(365)(8.33)CFECL3 \qquad (56)$$

where:

FECL3 = *per cent ferric chloride required by weight*

DFECL3 = yearly cost of ferric chloride

CFECL3 = cost of ferric chloride per pound

3.1.8 Sand Drying Beds

The routine for sand drying beds is contained in the portion of the program calculating costs. Vacuum filtration, elutriation, and incineration costs are set equal to zero, and the beds are designed on the basis of a loading factor *of 4.4 pounds of solids per square foot of bed per month.* This is a very arbitrary method, and in no way takes into account the properties of the sludge. The cost of $2.23 per square foot for construction also seems very high.

3.1.9 Sludge Incineration

Smith had difficulty obtaining accurate and plentiful data in which the capital and operating costs could be found separately, but managed to approximate the costs as follows:

$$\text{capital cost} = 1570(\text{pounds dry solids/day})^{.6} \quad (57)$$

$$\text{operating costs} = 16.1(\text{YTONS}) - 0.00009(\text{YTONS})^2 \quad (58)$$

where: capital cost = dollars

operating costs = dollars/year

YTONS = total number of tons of sludge solids incinerated per year assuming 70-75% moisture

Smith indicated dissatisfaction with this method, and predicted new work to correct the situation.

3.1.10 Cost Relationships

The very heart of this model is the cost analysis that predicts a total cost for treatment of the wastewater. The equations calculating the costs are based upon an Engineering News Record Construction Cost Index of 812, and a capital cost index (CCI) is input to the simulation to account for construction cost increases. The cost relationships were based upon an actual cost estimation of plants ranging from 0.25 to 100 MGD, and curves were fit to the data, based upon the major design parameters, such as the air blowers. Table I shows the capital cost relationships incorporated into the model.

Also input into the model are excess capacity factors for each process to allow for cleaning, repair, or storage. For example, in colder climates, excess capacity might be required for the sludge digesters if sand drying beds are to be utilized. For example, an excess capacity factor of 2 is recommended for the anaerobic digesters. This means that the design capacity calculated by the simulation is doubled, the size required is determined, and before the cost is predicted, the volume of the digester will be adjusted. The factors suggested are listed in Table III.

Table II summarizes the relations used to describe the operating costs for the plant units. The thickening and elutriation tanks are assumed to have only small costs associated with their operation, and are thus considered negligible. An unfortunate feature of these cost relationships is the fact that operating costs for drying beds are included in the operating costs of the anaerobic digester. If anaerobic digestion and sand beds are used, then the cost is hidden. This also says that the cost for the sand beds is independent of anything but the size of the digester.

TABLE I

Unit Process Capital Cost Relationships, after Smith (2)

Unit Process	Cost Equation (Dollars)
Preliminary Treatment	$14,700(Q_1)^{0.625}$
Primary Sedimentation	$13,400(APS) + 5200(APS)^{0.1}$
Aerator	$175,000(VAER) + 36,500(VAER)^{0.182}$
Air Blowers	$10,750 + 5,857(BSIZE)$
Final Settler	$12,600(AFS) + 5350/(AFS)^{0.126}$
Sludge Return Pumps	$3650 + 1125(Q_6)$
Control House	$40,000(Q_1)^{0.7}$
Digesters	$5000 + 1080(VDIG) + 10700(VDIG)^{0.128}$
Vacuum Filters	$12,800 + 372(AVF)$
Sludge Drying Beds	$2.23(ASB)$
Chlorine Contact Tank	$9000(Q_5)^{0.469}$
Plant Site	$4400(Q_{20})^{0.875}$
Sludge Thickeners	$(18.1 + 8.46/\exp(ATHM/13375))ATHM$
Sludge Elutriation Tanks	$(18.1 + 49.5/\exp(AE/6000))AE$

TABLE II

Unit Process Operating Cost Relationships, after Smith (2)

Unit Process	Operating Cost (Dollars/Yr.)
Preliminary Treatment	$500(Q_1) + 2150(Q_1)^{0.37}$
Primary Settling	$1000(APS) + 2500(APS)^{0.5}$
Aerator, final settler Blowers-excluding power	$10,000(VAER) + 14500(VAER)^{0.63}$
Digesters	$48(VDIG) + 540(VDIG)^{0.44}$
Digesters and sludge Drying Beds	$80(VDIG) + 900(VDIG)^{0.44}$
Vacuum Filter-excluding chemicals	$1500(Q_1) + 6450(Q_1)^{0.37}$

TABLE III

Excess Capacity Factors as Recommended by Smith, (2)

Unit Process	Excess Capacity Factor
Preliminary Treatment	1.0
Primary Settler	2.0
Aerator	1.2
Air Blowers	1.5
Final Settler	2.0
Sludge Return Pumps	2.0
Control House	1.0
Thickener	1.5
Digesters	2.0
Sludge Elutriation	1.5
Vacuum Filter	1.0
Incinerator	1.0
Sludge Drying Beds	1.0
Chlorination	1.0
Plant Site Preparation	1.0

No cost was assessed for heating the digesters, and similarly, no credit was given for the utilization of energy produced in the digester gas. The cost of ferric chloride used for sludge conditioning was taken as $.08/lb., and the cost of electricity was assumed to be $0.01/KWHR.

3.2 Deletion of Primary Sedimentation

Usually, all conventional biological treatment processes are preceded by preliminary sedimentation. Sand filters and trickling filters would tend to become clogged by the solids, and the additional solids would place an increased load upon the biota in the aeration tanks of the activated sludge process without sedimentation.

However, with the advent of the small package plant to serve small populations which previously could not afford biological treatment came extended aeration and contact stabilization, along with many other variations of the activated sludge process. Extended aeration provides long term aeration of the sewage for periods up to twenty-four hours, with or without final sedimentation. Long aeration periods allow for release of a well stabilized floc to the receiving water, and shorter aeration periods with final sedimentation and sludge return provide for a clear effluent and a higher solids content in the aerator. Contact stabilization in small plants provides for moderate aeration periods ranging from four to eight hours with recirculation of the sludge from the final settler which has been activated in an aerobic digester.

In larger sized plants, Setter et al. (18) reported on the modified aeration system employed at the Jamaica Sewage Treatment Plant, in the City of New York. The plant had a daily average flow of 35 MGD, and employed short to medium term aeration coupled with low and high mixed liquor suspended solids values and omission of primary sedimentation. Aeration periods of 2.0 - 2.5 hours, and of 4.0 - 4.5 hours were used for the short and long aeration periods, respectively, while 200 - 500 ppm, and 400 - 1200 ppm were values used for low and high suspended solids, respectively.

Average BOD reductions of 72.5% and suspended solids removals of 84.8% were reported in the effluent from the final settler for the entire 14 month testing period. Specifically, for the various combinations of detention time and solids loading, the BOD reductions were:

a) Low solids-Long aeration --------- 77.2%

b) High solids-Short aeration--------- 71.0%

c) Low solids-Short aeration --------- 69.9%

d) High solids-Long aeration --------- 76.6%

These experiments at the Jamaica plant indicated that it is feasible to omit primary sedimentation. However the result in this case is a somewhat lower BOD reduction as compared to the standard activated sludge process. The savings are in the primary settling tanks, smaller aeration tanks, and in a reduction of air to be supplied. The air required per pound of BOD removed is about twenty-two per cent lower, and the total air required is about one-third lower than a standard activated sludge plant. The Jamaica plant purchased only 7.2% of the gas energy used, and 6.8% of the electricity used. The remainder of the gas used was digester gas, and the remaining electricity used was generated using digester gas as fuel.

Studies by Heukelekian and Weisberg(9) on the causes of sludge bulking found that several steps taken to prevent sludge bulking would be aided by omitting primary treatment. They studied the relationships between bound water in the sludge and the sludge volume index.

It was found that sludges with high values for bound water had relatively high sludge volume indices. Additionally, when the sludges that bulked were subjected to continuous aeration, the bound water and the sludge volume index values both decreased. During this aeration period, the sludge color was observed to change from grey to golden brown, and the microbial colonies became dense and opaque.

The effect of chlorination upon sludge bulking was studied, and it was observed in a full scale plant that when the chlorination of the return sludge was discontinued, the sludge volume index increased progressively over a period of nine days. During the same time the chlorination was discontinued, the dissolved oxygen in the aerator increased from 0.8 mg/l to 3.0 mg/l. This indicates that the chlorine did not adversely affect the bacterial action in the sludge. The decrease in sludge volume index brought about by the chlorine was attributed to a physical release of bound water brought about by the chlorine.

The final experiments on the bulking of sludge were an attempt to determine the causes of the bulking, and under which conditions it occurs. The sludge was fed two types of sewage solids: soluble-colloidal solids obtained by centrifuging settled sewage; and solids obtained by settling sewage by gravity. It was reasoned that the soluble-colloidal solids would be representative of the solids passing through primary sedimentation, and that the settleable solids would represent the effect of eliminating primary treatment.

The sludge fed on the soluble-colloidal material demonstrated an increase in bound water and sludge volume index over the sludge that was not fed at all. The loading rates of 0.33, 0.5, and 1.0 pounds BOD per pound of mixed liquor solids were tested. The sample fed at the rate of 0.33 lbs BOD/lbs solids showed only slight increases, while the sample fed at 0.5 lbs BOD/lbs solids increased both bound water and SVI by a factor of 3.5; it did produce a clear supernatent however. The sample fed at 1.0 lbs BOD/

lbs solids showed increases only on the order of 2.6, but produced a turbid supernatent with a BOD of 45 ppm as compared to 10 ppm for the other two samples.

The sludge fed the settleable solids from sewage demonstrated a slight decrease in the bound water and SVI values, about a 20% decrease, and this held steady over the entire range of feed rates.

In summary, Heukelekian and Weisberg concluded that the elimination of primary sedimentation would allow the settleable solids to be incorporated into the activated sludge, thus increasing its settleability and would not materially increase the oxygen demand in the aerator. The active sludge solids fed on the settleable solids portion of raw sewage could not be made to bulk under any laboratory conditions. Chlorination was shown to be an aid in controlling the sludge volume index, and the dissolved oxygen level in the aerator or sludge was discounted as having any influence upon the bulking characteristics of the sludge.

3.3 Aerobic Digestion Studies

The sewage solids removed from wastewater by the treatment processes are highly organic in nature, and are very putrescent. It is usually desirable to stabilize these solids by a biological process before final disposal. By far the most popular process for biological stabilization up until now has been the anaerobic digestion process in which the organic material is first transformed into organic acids and then the acids are oxidized to carbon dioxide and methane. Volatile solids are reduced by 35-60%, and the remaining volatile solids are fairly stable and are not subject to rapid putrifaction.

The aerobic digestion process is intended to utilize the same microscopic biota present in the activated sludge process instead of developing a new ecology as is done in the anaerobic digestion process. The solids and bacteria are aerated for extended periods to promote full utilization of the organic food source, and then to allow endogenous respiration to reduce the microbial population and leave behind only the stable oxidized remnants of the cell structure.

The main intent of the digestion process is the reduction of volatile solids, which comprise 75 to 90 per cent of domestic activated sludge (6), and about 75 per cent for raw primary sludge (19). To establish the effect of aerobic digestion upon sewage sludge, many experiments have been carried out, mostly upon secondary biological sludges

Studies by Eckenfelder (6) found that the oxygen utilization of a pilot plant at Ridgewood, New Jersey, varied between 3 and 7 mg of oxygen per hour per gram of sludge, depending upon how long the sludge had been aerated. He also reported a mean oxidation rate of 5.2% per day at 21 to 24^{o}C. for the volatile solids. Also observed was the fact that nitrogen was solubilized during oxidation in approximately the same stoichiometric ratio to the amount of sludge oxidized.

Dreier (5) summarizes work done at the University of Wisconsin from 1959 to 1961 on the aerobic digestion process on a waste made up of one-third pre-treated meat packing waste, and two-thirds domestic waste. The studies were bench scale, carried out in four liter flasks inverted with holes in the bottom submerged partially in water baths to maintain constant temperatures. The feed sludge was maintained at a concentration close to 3.2%, and the volatile content was in the 70 - 80% range. The sludge was mixed in a ratio of 1.75/1.00 raw primary/waste activated on a dry solids basis.

Controlled variables were: detention time; loading rate (lbs volatile solids/ft^3/day); and temperature. The

The effect of temperature was found to be significant in tests using short detention times, but in tests using longer detention times, the results drew close together. The results on this were hard to show, for the longer detention times necessitated a lower loading rate since the solids content of the feed and the digester were regulated. The change in volatile solids after fifteen days was negligible at both 15 and 20°C. The detention time and the loading rate were not independent variables, as high loading rates forced low detention times.

The following seven conclusions were drawn:

1. Reduction of volatile solids is a function of detention time, noting that times over 15 days show a great slowdown.
2. Higher temperatures produce higher reductions of volatile solids.
3. In general, higher removals were obtained at lower loading rates.
4. Settling characteristics of sludges digested less than 30 days are poorer than for undigested sludge.
5. Sludges digested more than 5 days drain well
6. Supernatant characteristics are superior to anaerobic supernatant.
7. The sludge dried with no objectionable odor.

Further studies with higher loading rates indicated that volatile solids reductions were indeed significant after fifteen days, after digestion was found to be complete at lower loadings. For example: at a loading of 0.1125 lbs. volatile solids/cu ft/day, the reduction at fifteen days was 35%, and after thirty days the reduction was 53%. These studies found excellent correlation between the amount of solids destroyed and the sludge age. They propose the following relationship for determining the degree of volatile solids reduction:

$$\% \text{ Reduction} = 2.84 + 35.07 \text{ Log (sludge age)} \qquad (59)$$

This equation, fit through experimental data, was found to have a correlation coefficient of 0.78.

The use of sludge age as an independent parameter allows for the description of processes in which different solids concentrations are used in the feed, and for the practice of supernatant withdrawal, which would make use of a theoretical detention time difficult.

3.4 Sludge Dewatering on Sand Beds

Sludge drying beds are probably the oldest and most widespread method used for the dewatering of sewage sludges. Basically the beds are made up of a layer of sand overlaying a layer of gravel. Oftentimes an under-drain system will be provided to collect the drained liquid.

The dewatering process itself has been described as being composed of two mechanisms: drainage and drying. In drainage the liquid, or subnatant as it is sometimes called, flows down under the influence of gravity through the sludge and sand bed. It may be collected and returned to the plant, or just mixed with the effluent, or even be allowed to percolate into the ground. In the drying process, liquid evaporates from the surface of the sludge, and is lost to the atmosphere.

There have been several investigations into the design parameters to be used for sludge drying beds, but most of the earlier attempts were approached from the point of studying the performance of existing beds, and fitting curves through the data points to obtain empirical design relationships.

Skinner (20) came up with the equations:

$$\text{area per capita} = K \frac{(\text{average annual precipitation})\ (\text{suspended sewage solids: ppm})}{(\text{number of months in drying season})\ (\text{mean annual temp.: }^{o}\text{F.})\ (\text{mean wind velocity: mph})} \qquad (60)$$

where: K is a constant depending upon the treatment process. Glass covered beds required half of the calculated area.

Haseltine (8) plotted data received from the field and developed the following equation to predict sand bed performance:

$$Y = 0.157\ S_0 - 0.286 \qquad (61)$$

where: Y = solids loading ($kg/m^2/day$)

S_0 = solids content (%)

More recently, work concerning the mechanisms involved in the dewatering of sludges has been performed at the University of Massachusetts(17), (19), (2), (1). Studies have also been carried out on the drying of sludge by evaporation alone (13).

In the work on dewatering, the concept of specific resistance was used to describe the flow of a fluid through a porous media. The derivation started with the Darcy-Weishbach formulation for flow through a pipe, with the pores of the sand media comprising many small pipe-like conduits. The flow was observed to be slow, so laminar flow was assumed. Sanders (17) developed a mathematical model to describe the draining process.

$$t = \frac{\mu R_c \; S_0}{100(H_c)^{\sigma} \; (\sigma + 1)} \left[H^{\sigma+1} - \frac{\sigma+1}{\sigma} H_0 H^{\sigma} - H_0^{\sigma+1} + \frac{\sigma+1}{\sigma} H_0^{\sigma+1} \right] \quad (62)$$

The same equation is rearranged and stated as follows by Adrian and Nebiker (14)

$$t = \frac{\mu R_c \; S_0}{100\sigma \; (\sigma + 1) \; H_c{}^{\sigma}} \left[H_0{}^{\sigma+1} + \sigma H^{\sigma+1} - (\sigma + 1) \; H_0 H^{\sigma} \right] \quad (63)$$

where: R_c = Specific resistance at a head loss of $H_c (T^2 M^{-1})$

S_0 = Initial solids content (%)

H_c = Head loss at which R_c is measured (M)

H = Head at time t (M)

T = Time (sec.)

H_0 = Original head (M)

σ = coefficient of compressibility

μ = dynamic viscosity of the filtrate$(ML^{-1}T^{-1})$

A media factor, m, is introduced into the above equation to account for the difference in resistance to flow presente by differing porosities of the Buchner funnel and different grades of sand used in drying beds.

In the area of drying of sludges by evaporation, Nebiker(13) carried out studies on actual sludges drying in the open air without any drainage. It was found that water evaporating from the free surface of the sludge evaporated at a constant rate almost identical with the rate of evaporation of a free water surface. When the solids content of the cake rose to such a point that the upward movement of water by capillary action could not satisfy the evaporation potential of the surface, the drying rate entered what is called the falling rate stage. The following equation was proposed to describe the time required for drying by evaporation:

$$t = \frac{W_{ts}}{100\ A \cdot I_{s,c}} \left| U_o - U_{cr} + U_{cr} \cdot \ln(U_{cr}/U_t) \right| \qquad (64)$$

only when: $U_o \geq U_{cr} \geq U_t$

where: W_{ts} = weight of the total solids (grams)

A = surface area (Meters^2)

$I_{s,c}$ = constant rate drying intensity (kg/m^2/hr)

U_o = original water content (%)

U_{cr} = critical water content, when constant rate drying ceases (%).

U_t = water content at the end of drying (%)

t = time of drying (hours)

The water content at all times is calculated as:

$U = \frac{w}{1 - 0.01 \cdot w}$: w is the per cent solids

PART 4: MODEL THEORY

4.1 Deletion of Primary Treatment

The first work on the computer simulation model was to separate the functional groups from the FWPCA model into separate subroutines. Preliminary treatment, biological treatment, and final settling are contained in the first subroutine named BIOLOG. The second subroutine, PRIMARY, contains the model of primary treatment. If primary treatment is to be utilized, the subroutine PRIMARY is called by the subroutine BIOLOG. In the case that primary sedimentation is not to be used, it is only necessary to adjust the stream values at point #2 to be equal to the combination of the raw sewage mixed with the return stream, stream #9. Stream #8, the underflow from the primary settler also must be adjusted to zero. The model then progresses exactly as it was written by Smith(21).

4.2 Design Basis for Aerobic Digestion

The subroutines following PRIMARY, and BIOLOG, are: THICK, which describes the sludge thickening process; DIGEST, which describes the anaerobic digestion process and the elutriation process; VACUUM, which describes vacuum filtration; and SEWAGE, which is the main calling program also containing the input/output routine and the economic evaluations.

When aerobic digestion is to be used, the new subroutine, AEROBE, is called to replace THICK and DIGEST. Thickening and elutriation are not required when the sludge is aerobically digested because the sludge is thickened in the digestor by supernatant withdrawal, and the alkalinity of the aerobically digested sludge is considerably lower than it is for anaerobically digested sludge (12), (15), (5).

The reduction of volatile solids in the sludge is the prime purpose of the digestion process, and in this model, the empirical relation found by Lawton and Norman (10) will be used to predict the reduction of the volatile solids in the digestor:

$$\% \text{ reduction} = 2.84 + 35.07 * \log (\text{sludge age}) \qquad (65)$$

This equation was developed for the continuously fed, completely mixed system which in practical terms would mean daily feeding and daily withdrawal of supernatant and sludge if vacuum filters are used. If sand drying beds are employed the supernatant withdrawal would be daily, but sludge withdrawal would be on a less frequent schedule. A listing of the subroutine used to simulate the aerobic digestion is found in the Appendix.

The values for the required per cent volatile solids reduction and the total suspended solids to be maintained in the aerator are input to the program as is the per cent of solids escaping in the supernatant liquor. Then all of the stream vectors, BOD, TSS, VSS, SOC, etc. for the influent stream must be set equal to the values at station #10, which is the point of mixture for the primary and secondary sludge streams. The reason for this is that the thickening process is being omitted, and points #10 and #12 become the same point.

The main program, SEWAGE, now calls AEROBE instead of DIGEST, and before returning to the main program AEROBE sets all stream values at point #15 equal to the stream values at point #13 since elutriation is also by-passed.

The detention time in the digester is found by setting digestion time, DIGT, equal to seven days and solving for the volatile solids reduction. If the per cent volatile solids exceeds the required value input to the program, then the digestion time is incremented by adding one day and re-computing the volatile solids content. Seven days was used as an initial starting point because several investigators have found that low digestion periods yield a poorly draining sludge (12), (5), (15), (11). The same investigators also indicated that settleability also increased with longer detention times.

The volume of the supernatant withdrawn is found in two steps. First, the amount required to bring the thin influent sludge up to the solids content specified to be held in the aerator is found by the following equation:

Q(13) = Q(12)*TSS(12)/(solids*10,000.) (66)

where: solids = per cent solids held in the digester

Next the destruction of solids in the digester is taken into account by again increasing the withdrawn supernatant to compensate for the lower total solids:

$$Q(13) = Q(13)*TSSA/(SOLIDS*10{,}000.) \quad (67)$$

where: TSSA = the total suspended solids in the aerator after solids destruction is accounted for.

Now Q(13), the effluent volume has been found, and the volume of the supernatant is merely the difference between the influent volume and the effluent volume:

$$Q(11) = Q(12) - Q(13) \quad (68)$$

The qualities of the effluent and supernatant streams are now found. The total suspended solids in the supernatant is specified by the input variable, TSSLA, total suspended solids lost from the aerator. This value must be supplied and is a product of engineering and experimental experience. All solid species are assumed to be removed according to the same ratio, much in the same manner as in the thickening process set forth in Smith's model (21):

$$TSS(11) = TSSLA*SOLIDS*10{,}000. \quad (69)$$

$$VSS(11) = VSSA*TSSLA \quad (70)$$

$$SON(11) = SOLIDNA*TSSLA \quad (71)$$

$$SOP(11) = SOLIDPA*TSSLA \quad (72)$$

where: VSSA = volatile solids in the digester

SOLIDNA = solid nitrogen in the digester

SOLIDPA = solid phosphorous in the digester

The nitrogen and phosphorous converted to the soluble state is considered to be in direct ratio with the amount of volatile material destroyed. As the cellular material is broken down, the nitrogen and phosphorous are released to the environment. This agrees with the findings of Eckenfelder (6), in his studies on oxidation kinetics of biological sludges.

The volume of the digester is taken to be the product of the digestion time and the effluent volume plus the volume of one day's supernatant:

VOLDIG = DIGT*Q(13) + Q(11) (73)

Excess capacity for storage and cleaning is provided through the excess capacity factor input by the user.

The air requirements of the digester are found by providing for 15 to 20 cfm/1000 ft^3 of digester capacity. Loehr (11) found that the air required to maintain the solids in suspension was the limiting factor on the air supply:

AIRCFD = AIRCFD + VOLDIG*20.0*60.0*24.*1000./7.48 (74)

where: AIRCFD = cubic feet of air supplied daily

The blowers are sized by increasing capacity by the amount of air required by the digester:

BSIZE = AIRCFD/(24.*60.) (75)

where: BSIZE = Blower capacity in ft^3/min.

Finally the economic relationships in the main program, SEWAGE, are adjusted to determine the cost figures. The construction cost of the digester is taken to be the same as for a similar size activated sludge aerator:

$$CCOST(3) = 175{,}000(VOLDIG) + 36500(VOLDIG)^{0.182} \quad (76)$$

The operating cost of the digester is found by adding the size of the digester and the aerator and using the formula given by Smith (21) in the original model:

$$\text{dollars/year} = 10000.(VAER) + 14500(VAER)^{0.63} \quad (77)$$

The capital cost of the larger air blowers is found by the formula:

$$CCOST(4) = 10750 + 5857*BSIZE \quad (78)$$

The amortization costs are based upon a 25 year life and an interest rate of 4.5%. The construction cost index is set for the week of August 3-9, 1969 when the ENR INDEX was at 1275. The cost relations in the program were keyed to an INDEX value of 812, and a straight ratio is used to convert the capital costs to presently valid costs.

4.3 Design Basis for Sludge Drying Beds

The subroutine written to simulate sand drying beds utilizes equations #63 and 64 by Sanders (17) and Nebiker (13), respectively for the draining and the drying of sewage sludge. The program starts with a sludge depth of 1 centimeter, and increments the depth by 1 centimeter after the total dewatering and drying time is computed.

The draining time is assumed to be short enough so that the drying which goes on simultaneously will be insignificant. Therefore, the time required to drain to a certain solids content is determined first, and then the time required for the sludge to reach its final solids content is determined.

After the total time to dewater is determined,the amount of land required for the sand beds is determined by dividing the available drying time per year by the total time required to dewater each application of sludge plus a period for bed restoration. By assigning a land cost to the area required, and a construction cost for building the sand beds, it is possible to determine a capital cost for any depth of sludge application.

In trying to optimize the process, it is desirable to minimize the total capital and operating costs. This requires that an operating cost relationship be available. This is a weak point in the program. The original model by Smith (21) estimates sand bed operating costs as a function of the anaerobic digester size, regardless of depth of application. The model also uses a single design loading factor of 4.4 pounds of solids per square foot of bed per month.

In this investigation's attempt to describe more precisely the effect of various variables upon sand bed performance, it was necessary to come up with some kind of operating cost relationship. Therefore, the cost per ton as predicted by Smith's model (21) was used as a basis, assuming that the beds were loaded to a depth of about 20 centimeters, which is the common depth of operation. Realizing that the depth of filling should influence the operational cost, it was reasoned that a shallow depth of filling would mean a large number of cycles per year, and that this would necessarily increase labor costs. However, there is little or no information in the literature to describe how the cost varies. Therefore the cost is assumed to vary linearly with the number of drying cycles per year and with the total number of tons to be removed:

$$COSTO(13) = (AREA*70. + NOTONS * 3.)/NOTONS \qquad (79)$$

where: COSTO(13) = operating cost of the sand beds (dollars/ton)

AREA = Area of sand drying beds (acres)

NOTONS = number of tons removed per cycle

This yields an operating cost of about \$6.50/ton, depending upon the quality of the sludge applied to the beds.

In this manner, the total cost for dewatering the sludge can be optimized by computing the total cost for each depth and comparing it with the cost for the previous lower depth. When the new cost exceeds the previous cost, the iteration is stopped, and the optimum operating point is considered to be reached.

The qualities of the subnatant liquid are determined by assuming that a certain percentage of the solids in the sludge will escape and travel through the sand layer. The dissolved species are assumed to remain constant in concentration.

The computer version of the equations controlling draining and drying are as follows:

Draining -

Z = (XMU*SI*RC)/(100.*(HC**SIGMA)*SIGMA*(SIGMA + 1.))(80)

Z2 = (SIGMA + 1.)*HO*H**(SIGMA) (81)

T = (Z*(SIGMA + 1.) + SIGMA*H**(SIGMA + 1.)-Z2))/3600(82)

Drying -

TDRY = (WTS/(100.*SCI))-(UO*UC + UC*LOG(UC/UT)) (83)

where: Z and Z2 are dummy variables

XMU = dyanmic viscosity of filtrate (gm/cm-sec)

SI = original solids content (%)

RC = specific resistance (sec^2/gm)

SIGMA = coefficient of compressibility

HO = original depth of sludge and sand (cm)

HC = head loss corresponding to RC (cm)

H = total head at time T (cm)

T = time (hours)

TDRY = time (hours)

WTS = weight of the solids per unit area (kg/m^2)

UO = original moisture content

UT = final moisture content

UC = moisture content at the end of constant rate drying

SCI = constant drying rate of a free surface ($kg/m^2 hr$)

The Smith model (2), quotes a construction cost for sand drying beds at $2.23 per square foot, keyed to an <u>Engineering</u> <u>News</u> <u>Record</u> index of 812. This seemed very high, so a rough approximation for a sand drying bed was attempted, using the "Building Construction Cost Data" values to estimate cost (4). The following basic design was arrived at:

1 Sand Drying Bed = 1 acre: 43,560 sq ft

depth: 70 cm - 10" gravel
17.7" sand

Drain pipe: 4", 20' on center

1/2" Asphalt Liner

Concrete curbing

Installed costs:

36,300 ft^3 gravel @ $4.00/yd =	$ 5,390.00
64,100 ft^3 sand @ $4.50/yd =	10,700.00
840 ft concrete curbing @ $2.65/ft =	2,220.00
43,560 ft^2 asphalt liner @ $0.40/$yd^2$ =	17,400.00
2,310 ft underdrain @ $0.90/ft =	2,080.00
Total cost per acre:	$37,790.00
Total cost per sq ft	$ 0.87

This figure of \$0.87 per square foot does not include any distribution piping or valves, but it is hard to conceive of a cost exceeding \$1.00 per square foot with miscellaneous piping added in. In terms of the ENR Index of 1960 at 812, the cost to be input to the program would be about \$0.64. Also not included in the above cost is any type of equipment for the removal of the sludge. This type of equipment would very much influence the operating costs of the sand drying beds. Thus it is assumed that the beds are cleaned by hand.

Land costs for the sand beds are input to the model. The cost is assumed to be \$10,000/acre in the simulations in this investigation.

PART 5: RESULTS

5.1 Cost and Performance Comparisons in the Deletion of Primary Treatment

When Primary sedimentation is omitted from the standard activated sludge treatment scheme, certain cost benefits accrue from the elimination of the primary settler. However, this will also require modifications to the aerator, and also to the sludge handling processes. Table A1 in the Appendix shows a complete analysis of the standard activated sludge plant with anaerobic digestion, while Table A2 shows a complete analysis of the activated sludge plant when primary sedimentation is omitted, leaving all other variables constant.

These tables will show that the quantity of sludge to be digested or otherwise treated is much greater when primary sedimentation is omitted in this way, for there is not a concentrated primary sludge to be mixed with the more dilute biological sludge from the aerator. Instead, all the solids pass into the aerator and are predicted to be removed in the same concentration as a normal biological sludge.

Table IV shows a cost comparison between the standard activated sludge and the modification where primary treatment is omitted. The mixed liquor suspended solids values

TABLE IV

Comparison of Activated Sludge with and without Primary Treatment

Cost: Cents/1000 gallons
URSS = 3.0

FLOW (MGD)	Primary	Non-Primary	Savings (%)
	MLSS = 2000 mg/l		
1.0	24.12	23.94	0.75
5.0	15.46	15.56	-0.66
10.0	13.58	13.73	-1.10
25.0	11.88	12.08	-1.68
50.0	10.97	11.20	-2.10
100.0	10.29	10.53	-2.32
	MLSS = 3000 mg/l		
1.0	23.01	22.30	3.08
5.0	14.58	14.21	2.53
10.0	12.74	12.45	2.27
25.0	11.09	10.86	2.07
50.0	10.20	9.99	2.00
100.0	9.54	9.35	1.94
	MLSS = 4000 mg/l		
1.0	22.40	21.42	4.37
5.0	14.12	13.51	4.32
10.0	12.32	11.79	4.30
25.0	10.69	10.23	4.29
50.0	9.81	9.39	4.27
100.0	9.16	8.76	4.35

are held constant in both types, and the concentration of the sludge underflow is held constant. The figures show a slightly more expensive treatment cost for the scheme that omits primary treatment when the mixed liquor suspended solids value is held at 2000 mg/l. The cost difference falls in the range of 0.75% less expensive for small plants to 2.32% more expensive for larger plants. The computer simulation should not be expected to be accurate enough to make these differences significant. However, when the mixed liquor suspended solids held in the aerator is set at values of 3000 mg/l and 4000 mg/l, the simulation predicts that the omission of primary treatment will result in a savings of between 3.08% and 4.37%.

The above results were founde by assuming that the mixed liquor suspended solids level to be the same for both systems. However, commonly used designed parameters for activated sludge aerator loadings are based upon a ratio of pounds of BOD entering the aerator to the number of pounds of mixed liquor suspended solids in the aerator.

Primary treatment normally is figured to remove about 35% of the raw BOD. Since this is not the case when primary settling is deleted, the two systems would not be loaded equally with the same value of the mixed liquor suspended solids. A more valid comparison between the two systems could be obtained by adjusting the aerator loadings so that they were more closely equivalent. Since the system.

in which primary treatment is deleted receives about 50% higher BOD in its influent stream to the aerator, a simulation was made with the mixed liquor suspended solids for the non-primary system set at 150% of the value for the system employing primary treatment. The results of this simulation compared with the standard activated sludge treatment scheme are contained in Table V.

This comparison shows that omitting primary sedimentation can yield savings ranging from 7.53% to 11.01%. The higher the level of mixed liquor suspended solids held in the aerator, the greater are the savings. While it is not common to see plants with mixed liquor suspended solids as high as 6000 mg/l, values as high as 9000 mg/l were observed at the Lee, Massachusetts plant. A summary of operating data observed at the Lee plant is presented in Table X.

Another factor which should be taken into account when attempting to simulate a plant in which there is no primary sedimentation, is the effect of the solid which would otherwise be removed by sedimentation before entering the aerator. When the solids content of the sludge leaving the final settler is increased, as predicted by Heukelekian and Weisberg (9), the sludge handling facilities become less expensive to construct and operate. Table A3 in the Appendix shows a complete simulation when the ratio of sludge solids to aerator solids (URSS) is set at 6.0. In

TABLE V

Comparison of Activated Sludge With and Without Primary Treatment

Cost: Cents/1000 gallons
URSS = 3.0 in both cases

FLOW (MGD)	Primary	Non-Primary	Savings (%)
	MLSS = 2000 mg/l: Primary MLSS = 3000 mg/l: Non-Primary		
1.0	24.12	22.30	7.53
5.0	15.46	14.21	8.10
10.0	13.58	12.45	8.33
25.0	11.88	10.86	8.61
50.0	10.97	10.00	8.85
100.0	10.29	9.35	9.20
	MLSS = 3000 mg/l: Primary MLSS - 4500 mg/l: Non-Primary		
1.0	23.01	21.12	8.20
5.0	14.58	13.28	8.90
10.0	12.74	11.57	9.20
25.0	11.09	10.02	9.65
50.0	10.20	9.18	10.00
100.0	9.54	8.56	10.31
	MLSS = 4000 mg/l: Primary MLSS = 6000 mg/l: Non-Primary		
1.0	22.40	20.48	8.55
5.0	14.12	12.79	9.42
10.0	12.32	11.12	9.75
25.0	10.69	9.60	10.20
50.0	9.80	8.77	10.50
100.0	9.16	8.15	11.01

this simulation the mixed liquor suspended solids in the aerator was set at 3000/mg/l, which caused the secondary sludge to be predicted as being 1.8% solids. This is not impossible, as can be seen from the observations at Lee, Massachusetts, in Table X.

The savings resulting from this type of simulation are increased slightly from the case in which only the mixed liquor suspended solids are increased to balance the increase in BOD loading. As shown in Table VI, the costs are from 8.35% to 12.40% less than a corresponding standard activated sludge system.

TABLE VI

Comparison of Activated Sludge With and Without Primary Treatment

Cost: Cents/1000 gallons
URSS = 3.0 with Primary
URSS = 6.0 without Primary

FLOW (MGD)	Primary	Non-Primary	Savings (%)
	MLSS = 2000 mg/l: Primary MLSS = 3000 mg/l: Non-Primary		
1.0	24.12	22.11	8.35
5.0	15.46	14.05	9.10
10.0	13.58	12.29	9.50
25.0	11.88	10.70	9.90
50.0	10.97	9.84	10.60
100.0	10.29	9.19	10.65
	MLSS = 3000 mg/l: Primary MLSS = 4500 mg/l: Non-Primary		
1.0	23.01	20.92	9.10
5.0	14.58	13.11	10.10
10.0	12.74	11.41	10.40
25.0	11.09	9.86	11.10
50.0	10.20	9.02	11.50
100.0	9.54	8.40	11.90
	MLSS = 4000 mg/l: Primary MLSS = 6000 mg/l: Non-Primary		
1.0	22.40	20.32	9.30
5.0	14.12	12.63	10.55
10.0	12.32	10.95	11.10
25.0	10.69	9.44	11.75
50.0	9.81	8.61	12.25
100.0	9.16	7.99	12.40

5.2 Cost Comparisons Between Standard Anaerobic Digestion and Aerobic Digestion

The substitution of aerobic digestion for the more standard anaerobic digestion process seems to be a very real possibility as predicted by the computer simulation. Table A4, in the Appendix shows a complete simulation for an activated sludge plant utilizing aerobic digestion. With a 60% destruction of volatile solids, and maintaining 6% solids in the digester, the predicted costs come out very close to the predictions for anaerobic digestion as contained in Table A1. The cost for the system utlizing aerobic digestion is predicted to be 0.22% less expensive. This is a small difference when it is realized that the cost predictions are only results obtained by a curve of best fit through experimental data. The small cost advantage might be well worth while when the simplicity of the system is examined.

Several other factors might also come in to play. The relative simplicity of construction of an aerobic digester as compared to an anaerobic digester might make it desirable to construct smaller, subdivided aerators for the digester, and perhaps use a smaller excess capacity factor than is used for an anaerobic digester.

A closer look at the predicted cost figures for this entire plant shows that the cost of the aerator for the digester is between 50% and 60% of the cost of the anaerobic digester and its required conditioning processes of thickening and elutriation. This includes the extra blower capacity required by the aerobic digester. However, the cost for vacuum filtration and incineration increase because of the increased volume of sludge to be disposed of, and the capital costs for the vacuum filter and incinerator are more than double the capital cost of the aerobic digester. Changes in digester performance such as the level of solids and the per cent volatile solids destruction have a great effect upon the operation and size of the vacuum filter and incinerator.

Table VII shows just how close the costs for a plant with aerobic digestion compare to the costs for a plant with anaerobic digestion. The range of cost difference is between .42% cheaper for aerobic digestion, and 2.04% more expensive for flows ranging from 1 to 100 million gallons per day. This comparison is again for a volatile solids destruction of 60% and a solids content of 6%.

Table VIII shows the effect of different operating conditions in the aerobic digester upon total treatment costs for a 10 MGD plant. When the solids destruction or the solids level in the aerator declines, the costs for treatment increase. This is due to the increased hydraulic load upon the vacuum filter. The incineration costs also rise with increased water content of the vacuum filter cake, which is a function of the solids content of the sludge applied to the vacuum filter. The rate of volatile solids reduction can be controlled readily by adjusting the detention time in the aerator. However, the solids content of the aerator is a factor which may not be as readily controlled,although settleability usually is found to increase with detention time, and it is settleability which determines the amount of concentration which can be accomplished.

TABLE VII

Comparison of Anaerobic Digestion and Aerobic Digestion for Activated Sludge

Costs: Cents/1000 gallons
MLSS = 2000 mg/l

VOLAT = 60% SOLIDS = 6.0%

FLOW (MGD)	Anaerobic	Aerobic	Savings (%)
1.0	24.12	24.10	<0.1
5.0	15.46	15.41	0.32
10.0	13.58	13.55	0.22
25.0	11.88	11.93	-0.42
50.0	10.97	11.11	-1.27
100.0	10.29	10.50	-2.04

TABLE VIII

The Effect of Varying Volatile Solids Reduction and Aerator Solids Content on the Cost Comparisons Between Aerobic and Anaerobic Digestion

Flow = 10 MGD
MLSS = 2000 mg/l

VOLAT (%)	SOLIDS (%)	ANAEROBIC ¢/1000 gal	AEROBIC ¢/1000 gal	SAVINGS (%)
60.0	6.0	13.58	13.55	0.22
60.0	5.0	13.58	13.89	-2.28
60.0	4.0	13.58	14.18	-4.40
50.0	5.0	13.58	14.01	-3.16

5.3 Cost Comparisons Between Vacuum Filtration and Sand Drying Beds as a Means of Sludge Dewatering

Utilizing the computer simulation for gravity dewatering as a part of the standard activated sludge process, cost figures were obtained for comparison with vacuum filter and incinerator operations for sludge disposal. Table A5 in the Appendix shows a complete simulation for activated sludge with sand bed dewatering. By comparing these costs with the results shown in Table A1, which shows predictions for plants with vacuum filtration and incineration, an idea of the relative economies of sand bed dewatering can be gained. The results in Table A5 assume 365 days a year available for drying, a value which is not universally acceptable. However, to assume any other value would be to localize the simulation to particular areas. This is valuable when applying the model to particular cases, but is too specific for general discussion. When the relative magnitudes of the costs are discussed, the effects of differing drying periods will be indicated.

The total treatment costs for the system utilizing sand beds are shown to be about 22% less than the costs for mechanical dewatering and incineration. The drastic effect of the sand beds can be seen by comparing the amortization and operating costs (¢/1000 gal). The cost of the vacuum filtration and incineration was 3.703¢/1000 gal while the cost for the sand drying beds was 1.203¢/1000 gal.

Two factors come into play when one wishes to consider the effect of fewer than 365 drying days per year: the size of the sand beds required to accomplish drying in a shorter period; and the extra storage required in the digester or other tank to hold the sludge during periods when it cannot be applied to the beds.

The area required to dry a given amount of sludge varies linearly with the number of drying days available. Sludge which requires two acres to dry in 365 days will require four acres to dry in only half that time. However, the effect of the extra storage required depends upon whether the decrease in the number of available days is due to inclement weather such as rain, or due to a longer period or season such as winter when no drying can be accomplished for long periods of time.

In the case in which available drying days are reduced because of rain, the excess capacity factor built into the plant will normally be sufficient to smooth over the irregularities in the weather. Thus the only extra cost incurred would be in the larger sand beds.

In the case in which there exists a long season of no drying days, extra storage must be provided in the plant to carry over non-drying season. Thus in this case extra costs are incurred thorugh both larger sand beds, and in provision of extra sludge storage facilities.

The excess capacity factor normally applied to the anaerobic digester is 2.0, meaning that for a conservative design, twice the theoretical digestion volume is provided. This would easily smooth over the effects of reduced drying days due to rain. Thus if only 182 drying days were available due to precipitation, roughly twice as many sand beds would be required, and the dewatering costs would roughly double to 2.4¢/1000 gal, still significantly less than the cost for the same size plant utilizing vacuum filtration and incineration.

The effects for cases in which the drying season is restricted by longer periods of inclement weather require analysis for each individual case. However, the cost differences are so large, the author feels that most plants in this country could economically utilize sand bed dewatering.

The effects of scale of the plant on the two methods of dewatering are shown in Table IX. In both cases, the sludge is digested by the anaerobic digestion process.

TABLE IX

Comparison of Sand Bed Dewatering, and Vacuum Filtration Plus Incineration; As Reflected In Total Treatment Costs

Costs: Cents/1000 gal

FLOW (MGD)	Vacuum Filter and Incineration	Sand Drying Beds	Savings (%)
1.0	24.12	17.27	28.4
5.0	15.46	11.89	23.0
10.0	13.58	10.92	19.5
25.0	11.88	10.11	14.9
50.0	10.97	9.78	10.8
100.0	10.29	9.59	6.8

5.4 Performance Comparisons with an Actual Treatment Plant

The Water Pollution Control Plant at Lee, Massachusetts, was used to compare the predictions of the computer model with actual performance data. The treatment processes are shown in flow diagram form in Figure 2. Primary treatment is not employed, and aerobic digestion is used instead of anaerobic digestion. Sand drying beds are also used, although no performance data was taken from these. The specific resistance and coefficient of compressibility were determined for the aerobic digester sludge, and these values can be used to characterize the performance of the sludge on the drying beds.

The treatment plant utilizes diffused air for its aeration tanks and aerated grit chamber. The plant is new, and is rated at roughly one million gallons per day, and is presently treating only about three hundred thousand gallons per day. One half of the aeration tank capacity is being utilized, and all of the secondary settler capacity is being used.

The operation of the plant is such that very high solids levels are maintained in the aerator, with total sludge recirculation being employed. The aerators of the digester are fed about once a week, after the supernatant is withdrawn. Daily or regular supernatant withdrawal is

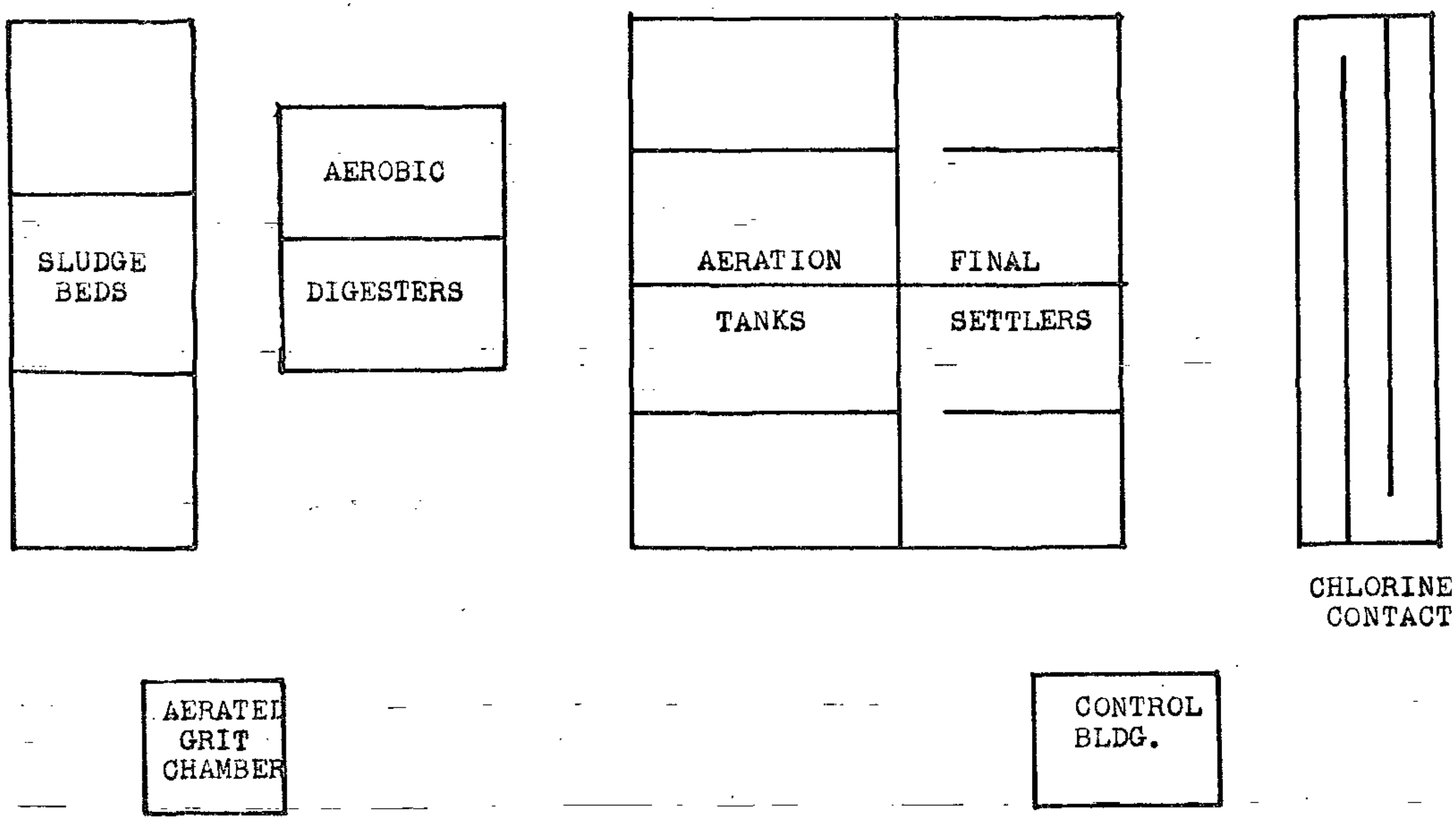

Figure 2.

WATER POLLUTION CONTROL PLANT, LEE, MASSACHUSETTS not practiced at the moment, probably because of the excess capacity now available.

The experimental results of samples from the Lee treatment plant are found in Table X. Five day BOD tests on the raw sewage, the plant effluent, and the digester supernatant were run on samples composited from samples collected at 8 A.M., 12 Noon, and 4 P.M., and run according to the methods in Standard Methods (22). The following solids tests were run on the raw sewage, final effluent, aerator liquor, raw sludge, and digester liquor; residue upon evaporation; and residue upon ignition at 600^{o}C. Both tests were performed on the solids retained on No. 2 Whatman Filter Paper. The solids tests were taken from grab samples, and dried at 103^{o}C to constant weight at the treatment plant and then brought to the University of Massachusetts Environmental Engineering Laboratories for ignition at 600^{o}C.

The treatment plant achieved an average of over 96% BOD removal as indicated by the laboratory results, thus leaving less than half of the required 13.0 mg/l of BOD in the effluent as predicted by the computer model.

The results of the solids tests on the raw sludge, the aerator liquor, and the digester liquor show an interesting pattern. On the first day of testing, one digester was being emptied and cleaned. The digester in

TABLE X

Experimental Results from the Lee, Massachusetts Treatment Plant

TEST POINT	SUSPENDED SOLIDS (mg/l)	VOLATILE SUSPENDED SOLIDS (mg/l)	VOLATILITY (%)	BOD (mg/l)
Raw Sewage				
#1	177.0	147.0	83.0	206.0
#2	103.0	81.0	78.5	198.0
#3	157.0	129.0	82.1	217.0
Average:	146.0	119.0	81.2	207.0
Final Effluent				
#1	10.60	7.11	67.0	4.12
#2	11.45	8.41	80.7	6.65
#3	12.10	9.70	80.0	8.60
Average:	11.38	8.40	75.9	6.46
Aerator Liquor				
#1	9000.0	4830.0	53.7	--
#2	11200.0	6300.0	61.7	--
#3	8050.0	4100.0	50.9	--
Average:	9417.0	5077.0	55.4	--

TABLE X (Continued)

Experimental Results from the Lee, Massachusetts Treatment Plant

TEST POINT	SUSPENDED SOLIDS(mg/l)	VOLATILE SUSPENDED SOLIDS (mg/l)	VOLATILITY (%)	BOD (mg/l)
Raw Sludge				
#1	31500.0	16500.0	52.3	--
#2	46700.0	24500.0	52.5	--
#3	18750.0	9200.0	49.0	--
Average:	32317.0	16733.0	51.3	--
Digester Liquor				
#1	47700.0	12100.0	25.4	--
#2	65100.0	16150.0	24.8	--
#3	20700.0	8300.0	40.0	--
Average:	44500.0	12183.0	27.4	--
Digester Supernatant				
#1	372.0	102.0	27.2	13.3
#2	783.0	240.0	30.6	17.9
#3	632.0	183.0	29.0	24.0
Average	596.0	175.0	28.9	18.4

use was not being fed, and the solids in the aerator were being totally recirculated. Five days later, when the second set of samples were evaluated, the total recirculation of aerator sludge was still being carried out, and the digester in use was decant thickened several times. This is reflected in the increased solids content of the aerator, raw sludge, and digester liquor.

The increase in the solids content of the raw sludge is attributed to two factors: The increased solids loading of the aerator; and the increased age of the mixed liquor suspended solids. Heukelekian in his studies on sludge bulking(8) found that the higher the sludge age, the better the settling characteristics. During this period, the aerator solids increased approximately 22% while the solids content of the raw sludge increased approximately 50%. Thus a part of the increase in the raw sludge solids content can be attributed to a higher mixed liquor suspended solids level, and a part to the increase in sludge age in which the sludge floc loses to some degree the gelatinous sheath which retains large amounts of water.

The next week, when the empty digester was being returned to service after cleaning, the solids content of the aerator predictably dropped, and with it the solids content of the raw sludge. The solids content of both digesters dropped as both were equalized and fed raw sludge.

The digesters are now being operated more like batch processes than continuous processes. Feeding is intermittent and supernatant withdrawal is infrequent. It is doubtful if this type operation can be maintained as the design flow of the plant is approached.

The dewatering characteristics of the aerobic sludge were determined using the specific resistance and coefficient of compressibility determinations as outlined by: Sanders, Adrian, Nebiker et al. (17), (14), (2), (1). The Buchner Funnel filter technique was used on two samples of aerobic sludge. Each sample of sludge was tested at four vacuum pressures, and the results were analyzed using a computer program developed at the University of Massachusetts under the sludge dewatering research programs. A complete sample of these tests appears in Appendix IV. Table XI shows a summary of the computer analysis of the data.

In each sample, the first coefficient calculated was lower than the ones using more data points. This could be due to a stray point in the data, but it did occur in both sets of experiments. The specific resistance values obtained are somewhat lower than the values reported by Nebiker, Sanders, and Adrian (19) of 4.8×10^{10} and 2.1×10^{10} sec^2/gm at 15 in Hg. for an anaerobically digested sludge from an activated sludge plant. Values

TABLE XI

Specific Resistance and Coefficient of Compressibility data Aerobic Sludge from Lee, Massachusetts

RUN	VACUUM (cm.Hg.)	SPECIFIC RESISTANCE (sec^2/gm)	COEFFICIENT* OF COMPRESSIBILITY
		SAMPLE #1	
1	24.9	$.8321 \times 10^9$	--
2	39.9	1.105×10^9	--
3	10.1	$.3069 \times 10^9$	--
4	59.9	1.804×10^9	.9712
		SAMPLE # 2	
1	25.0	$.8632 \times 10^9$	--
2	40.2	1.223×10^9	--
3	9.95	$.2997 \times 10^9$	--
4	60.0	1.877×10^9	.9753

*Note: The values for coefficient of compressibility are for the entire experimental test of four runs on each sample.

of 1.105 x 10^9 and 1.223 x 10^9 were obtained for the aerobically digested sludge in about the same vacuum range. These values represent a decrease in specific resistance by factors of about 17 to 43. This would materially effect the drainage time of the sludge, and therefore, lower the total dewatering time significantly.

A more significant difference lies in the coefficients of compressibility. Nebiker et al. (19) reported values of 0.63 and 0.64. The coefficients for the aerobic sludge were .9712 and .9753.

In comparison with the simulation of the Lee plant, the following parameters can be compared: BOD of the effluent; Volume of the aerator; suspended solids values in the aerator, raw sludge, and aerobic digester; and the volatile solids reduction in the aerobic digester.

The computer program predicts a final BOD of not more than 13 mg/l. The Lee treatment plant now is achieving a better than 96% BOD removal with an average of only 6.46 mg/l of BOD remaining. The volume of the Lee aerators is 750,000 gallons, much greater than the 231,000 gallons that the computer program predicts, and this may help explain the greater BOD removal being achieved at Lee.

The computer simulation varies the suspended solids content of the aerator according to the desire of the person using it, but it is interesting to note that the aerator at Lee has had suspended solids values ranging from 8,050 - 11,200 mg/l during the testing period, and had no problem with sludge bulking or excess solids in the effluent.

The suspended solids content of the raw sludge from the final settler varied from 18,750 - 31,500 mg/l, the lower value being observed when the solids content of the aerator was dropped by feeding the digester. The higher value was observed during the same period that high suspended solids values were observed in the aerator.

This is in accord with the findings of Heukelekian et al. (9) which showed that the greater the sludge age, the better is its settleability and compactability. The Lee results bear out the premise of greater solids content in the raw sludge when primary treatment is deleted.

The suspended solids values for the digester are input to the computer program, with a popular value being 6%. This value was observed at Lee just prior to feeding the digester. At the other times, digester feeding and the lack of supernatant withdrawal gave lower values for the digester solids content. The operators say there is no difficulty in maintaining high solids values,but that it is not now necessary at Lee because of the low flow conditions.

The volatility of the sludge entering the aerobic digester is about 51.3% and the volatility of the sludge leaving the digester is about 27.4%. This corresponds to a volatile solids reduction of 46.7%, somewhat below the 60% reduction called for in the computer simulation. However, the 46.7% is an average composited from values obtained when the digester had been in service for long periods yielding long detention times, and of periods when the digester had been heavily fed in the previous week, yielding shorter detention times. The highest solids reduction was observed to be over 52%. This value would probably have been greater if regular supernatant withdrawal had been practiced. This greatly increases the sludge age, which has been shown to be the determining factor in volatile solids reduction. Supernatant withdrawal is not practiced at this time.

Also, the volatile solids content of the raw sludge at Lee averaged only 51.3%, while the computer simulation predicts an average of about 85%. This may be due to the very long detention time in the aerator at Lee, 16 hours, and to the practice of total sludge recirculation. The digesters are fed infrequently, and high values of mixed liquor suspended solids are maintained in the aerator. The average value found during the testing period was 9,417 mg/l, which is very much greater than is normally expected. This practice of very high solids content in the aerator, and total sludge recirculation means that the sludge age in the aerator will be very high and that, in effect, the aerator becomes a digester for the sludge, greatly reducing the volatility before it enters the two separate digesters. Therefore, in actual practice the sludge volatility is reduced from about 85% to 27.4% as it travels through the aerator and digester. This corresponds to approximately a 67% reduction in total volatile solids.

In the comparison of the costs associated with the actual contract for the Lee plant, and the costs predicted by the computer simulation, very large discrepancies are apparent. The total contract cost of the Lee plant was $790,000, of which $123,000 was for general development not directly attributable to the cost of the plant. Therefore the total construction cost of the plant was $667,000. The computer simulation predicts a cost of $417,200 for a plant with the flow pattern and size of the Lee plant. This is 37.4% less than the actual contract cost of the plant as constructed.

Table XII shows the original engineer's estimate (23) and a pro-rata increase to take into account that the engineer's estimate was for $745,000, while the total contract cost was $790,000 including the general development work.

Table XIII shows a summary of the construction costs as predicted by the computer simulation. Comparing Table XII and Table XIII, it is seen that the cost categories are somewhat dissimilar. Table XIV is the result of an attempt to match the information into the same categories. In the engineer's report (23), the cost of the air blowers is included in the control house cost. Thus the costs for the control house and the air blowers will be combined in the simulation data. In the computer program,

TABLE XII

Engineers Construction Cost Estimates for Lee, Massachusetts Sewage Treatment Plant

UNIT	COST ($)	
	Original	Adjusted
Preliminary Treatment	$ 31,600.00	$ 33,600.00
Aeration Tank	166,000.00	178,300.00
Final Settler	83,300.00	89,800.00
Chlorination Tank	11,400.00	12,600.00
Digester	31,300.00	33,900.00
Control House	138,000.00	148,400.00
Outside Piping	54,000.00	58,300.00
Landscaping	27,000.00	29,300.00
Electrical Work	70,000.00	75,500.00
Drying Beds	8,000.00	8,800.00
	$622,000.00	$664,500.00

TABLE XIII

Unit Costs for Lee, Massachusetts Sewage Treatment Plant as Predicted by Computer Simulation

UNIT	COST ($)
Preliminary Treatment	$ 25,590.00
Aeration Tanks	121,100.00
Air Blowers	35,260.00
Final Settler	41,670.00
Sludge Pumps	10,070.00
Control House	69,620.00
Digester	66,220.00
Chlorination	15,660.00
Site Development	7,658.00
Drying Beds	24,440.00
	$417,200.00

TABLE XIV

Engineer's Estimated Costs, and Computer Simulation Costs Consolidated to the same Categories

UNIT	COST ($)	
	Engineer's Estimate	Computer Simulation
Preliminary Treatment	33,600.00	25,590.00
Aeration Tank	219,000.00	121,100.00
Final Settler	105,100.00	41,670.00
Digester	39,700.00	66,220.00
Control House	148,400.00	104,880.00
Chlorination	14,700.00	15,660.00
Drying Beds	8,000.00	24,440.00
Sludge Pumps	-------a	10,070.00
Site Development	29,300.00	7,658.00
Electrical Work	70,000.00	-------a
	$667,800.00	$417,200.00

a) Note: These figures are included in other categories the piping costs are included in the associated process unit costs, thus the outside piping costs from the engineer's report (23) will be divided between the aerator, final settler, chlorine contact chamber, and the digester. The cost will be proportioned according to relative costs of the individual units.

Checking the design parameters of the computer program with the design of the Lee plant, the major differences are found to lie in the size of the aerator and the air blowers. Smaller differences lie in the size of the digester and the final settler.

The Lee plant was designed on the basis of an extended aeration type process, and the engineers provided for a 16 1/2 hour detention time in the aerator at design flow. This required an aerator size of 0.675 million gallons. The computer simulation calls for a detention time of approximately 5 1/2 hours with an aerator size of only 0.237 million gallons. This represents almost three times the size aerator at the Lee plant as compared to the simulation.

Using the excess capacity factors to design the larger aeration tanks in the simulation, and also for increasing the associated blower size, it was found that \$143,342 was added to the aerator cost, and approximately \$64,800 was added to the cost of the air blowers.

These two changes increase the total plant cost by $208,142, and bring the total plant cost to $625,342. This represents only a 6.4% difference from the actual contract cost.

The cost differences represented by the slightly larger final settler at Lee, and the slightly larger digester predicted by the computer program just about exactly cancel each other. Thus, if the major design differences are taken into account, the cost of the actual plant can be readily verified by the computer program. It is the design engineer's responsibility to decide basic parameters, and these can be supplied to the simulation in order to ensure an accurate prediction.

From the experimental results taken from the Lee plant, it seems likely that such a long design detention period in the aerator is not necessary, however, this was one of the first plants of this type built on a large scale, and should serve as a model for future work. On the average, 3.42% of the raw BOD was all that was left in the effluent, and this was measured by a five-day BOD test. Most secondary activated sludge type plants achieve about 90 - 92% BOD removal. Remaining BOD removals require increasingly longer detention times.

The results of the solids determinations at Lee also suggest that the aerator is not being operated as an extended aeration plant in the usual sense of the word. The mixed liquor suspended solids levels in the aerator during the testing period averaged 9,417 mg/l, with almost total sludge recirculation. The digesters were only fed occasionally. This level is very much greater than the 2000 - 3000 mg/l level normally expected in standard activated sludge, and is certainly greater than the levels expected in extended aeration plants in which much of the stabilized solids escape to the receiving waters. The solids at Lee are very much stabilized in the aerator even before they are fed to the digesters.

The operating costs of the Lee plant were obtained from the town report, and from the Lee Sewer Commission. The budget for the first year of operation was approximately $29,500 and the budget for the second year's operation, 1969, was about $33,000. During the second year, 1969, an average of $342,000 gallons of sewage was treated daily. This works out to about 23.2¢/1000 gallons. Of this amount, all except $5,807, representing the cost of power is a fixed cost. The manpower and other maintenance costs are the same now as they will be when the plant is running at full capacity.

Using the projected cost of electricity at full capacity, $13,300/year, the operating cost will be about 9.9¢/1000 gallons. This is above the projected cost of 5.46¢/1000 gallons predicted by the computer program. However, if the large size aerator and blowers are taken into account, the projected operating cost would be about 8.6¢/1000 gallons,about 20% lower than the actual budgeted amount at Lee. However, it is interesting to note that the budget for the Lee treatment plant includes $5100 for Sewer Commissioner's salaries and general commission expenses. This represents about 17% of the 1968 budget, and this share should decrease as more sewage is treated.

It is very difficult to predict the operating costs of a treatment plant in general, for each owner or town has different ideas on salary levels for operators, the number of operators to be hired, the laboratory facilities to be provided, etc. These decisions are unique to each town or engineering firm. Thus an operating cost prediction should be an attempt to give a close approximation, but should not be expected to be very accurate.

PART 6: CONCLUSIONS.

6.1 Deletion of Primary Treatment

The computer simulation has shown that deleting primary treatment from the activated sludge process can yield a substantial savings in total treatment costs for almost all treatment plant sizes. Factors found to be important in this type design are:

a) the savings from elimination of a primary settler

b) increased cost due to a larger aerator

c) mixed liquor suspended solids values must be adjusted to equalize BOD loadings

d) the final sludge is considerably thicker than normal activated sludge, producing savings in the sludge handling group of processes.

Total treatment cost savings were predicted to be between 8.35 - 12.40%.

6.2 Aerobic Sludge Digestion

The simulation of the aerobic sludge digester showed that aerobic digestion can vary from slightly less expensive than anaerobic digestion, 0.22% @ 1.0 MGD, to slightly more expensive, 2.04% @ 100 MGD. The very small differences predicted by the simulation probably should be taken to mean that the two processes are very close in total cost, and other factors should be used to determine which process should be used. This conslusion was based upon the use of vacuum filtration and incineration for ultimate sludge disposal. If sand drying beds are used for final dewatering, the cost figures predict a substantial savings for the aerobic digester, for the digester itself is very much less expensive than the anaerobic digester. However, the vacuum filter is very sensitive to changes in solids and liquid loading rates, and the anaerobic process reduces the solids to a greater degree than the aerobic process. The aerobic digester costs only 42.4% as much as the anaerobic digester and its auxiliary processes of thickening and elutriation.

Aerobically digested sludge dewaters similarly to anaerobically digested sludge on drying beds as determined by specific resistance tests. However, the larger coefficient of compressibility for aerobically digested sludge compared to anaerobically digested sludge means the aerobically digested sludge would dewater somewhat faster.

6.3 Sand Drying Beds for Sludge Dewatering

The computer simulation showed that sludge dewatering on sand drying beds can yield significant savings as compared to the vacuum filter and incineration method of sludge disposal. Weather is a big factor in the costs of sand drying beds, and if 365 days of drying weather is assumed, the cost savings predicted range from 28.4% for a 1 MGD plant down to 6.8% for a 100 MGD plant. If only 182 drying days are assumed, the cost savings for a 10 MGD plant would still amount to about 7.5% over the same plant with vacuum filters and incineration.

The variables of weather, sludge characteristics and cost of construction are input to the simulation, and an optimum size of drying bed is derived.

The savings effected by sand drying beds can be important when deciding on the digestion process. Aerobically digested sludge tends to be more voluminous and consequently more expensive to dewater than anaerobically digested sludge having the same specific resistance, coefficient of compressibility and solids content for the simple reason that anaerobically digested sludge has more of its original solids destroyed, and therefore produces a slightly smaller volume of sludge to be dewatered. The cost of the sludge dewatering is more than twice as expensive as the actual digestion process, and is thus very sensitive to the differences in digester performance.

Sand drying beds on the other hand are less sensitive to variation in sludge quality, and can make aerobic digestion very attractive from a cost viewpoint.

PART 7: RECOMMENDATIONS:

7.1 Recommendations for Future Work

Suggested areas for future work include parts of the simulation now considered to be weak. These areas are:

a) More scientific evaluation of the elutriation process is needed.

b) Vacuum filter loading parameters which take into account factors such as solids content and sludge characteristics are needed.

c) More sand drying bed construction cost data would greatly improve the simulation.

d) Accurate sand drying bed operating data is sorely needed to improve the drying bed simulation.

e) A more scientific approach to the sludge thickener design would improve the engineering soundness of the model.

APPENDIX I

Complete Simulation Data Tables

TABLE A1

Standard Activated Sludge Process with Anaerobic Digestion and Vacuum Filtration

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISSOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	10.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	2000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

Table A1, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	10.318	.1160E+03	.1418E+03	.4021E+02	.1036E+02
	DIS		.4241E+02	.5874E+02	.1100E+02	.2200E+02
2	SOL	10.305	.5809E+02	.7098E+02	.2013E+02	.5436E+01
	DIS		.4241E+02	.5874E+02	.1100E+02	.2200E+02
5	SOL	10.111	.8903E+01	.9285E+01	.4261E+01	.1401E+01
	DIS		.1298E+02	.3695E+01	.1100E+02	.2347E+02
7	SOL	.195	.2150E+04	.2096E+04	.1029E+04	.1371E+03
	DIS		.1298E+02	.3695E+01	.1100E+02	.2347E+02
8	SOL	.013	.4641E+05	.5672E+05	.1608E+05	.4343E+04
	DIS		.4241E+02	.5874E+02	.1100E+02	.2200E+02
9	SOL	.320	.4611E+03	.1895E+03	.3598E+03	.3772E+02
	DIS		.2398E+02	.2427E+02	.1100E+02	.1160E+03
10	SOL	.208	.4900E+04	0	.1965E+04	.3985E+03
	DIS		.1481E+02	0	.1100E+02	.2338E+02
11	SOL	.170	.3000E+03	0	.1203E+03	.2440E+02
	DIS		.1481E+02	0	.1100E+02	.2338E+02
12	SOL	.038	.2541E+05	0	.1018E+05	.2066E+04
	DIS		.1481E+02	0	.1100E+02	.2338E+02
13	SOL	.038	.1039E+05	0	.1018E+05	.8525E+03
	DIS		.9842E+02	0	.1100E+02	.8121E+03
14	SOL	.138	.6947E+03	0	.6809E+03	.5700E+02
	DIS		.3434E+02	0	.1100E+02	.2206E+03
15	SOL	.014	.2080E+05	0	.2039E+05	.1707E+04
	DIS		.3434E+02	0	.1100E+02	.2206E+03
16	SOL	.012	.6934E+02	0	.6796E+02	.5689E+01
	DIS		.3434E+02	0	.1100E+02	.2206E+03
20	SOL	10.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

Table A1, Continued

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	10.318	.2154E+01	.3912E+02	.2471E+03	.2863E+03
	DIS		.4860E+01	.5000E+03		
2	SOL	10.305	.1078E+01	.1958E+02	.1237E+03	.1433E+03
	DIS		.4860E+01	.5000E+03		
5	SOL	10.111	.8903E-01	.3532E+01	.1692E+02	.2045E+02
	DIS		.5444E+01	.5000E+03		
7	SOL	.195	.2150E+02	.8531E+03	.4086E+04	.4939E+04
	DIS		.5444E+01	.5000E+03		
8	SOL	.013	.8617E+03	.1565E+05	.9886E+05	.1145E+06
	DIS		.4860E+01	.5000E+03		
9	SOL	.320	.6976E+01	.3246E+03	.9223E+03	.1247E+04
	DIS		.3177E+02	.5000E+03		
10	SOL	.208	.7371E+02	.1772E+04	.9801E+04	.1157E+05
	DIS		.5408E+01	.5000E+03		
11	SOL	.170	.4512E+01	.1085E+03	.6000E+03	.7085E+03
	DIS		.5408E+01	.5000E+03		
12	SOL	.038	.3821E+03	.9188E+04	.5081E+05	.6000E+05
	DIS		.5408E+01	.5000E+03		
13	SOL	.038	.1577E+03	.9188E+04	.2078E+05	.2997E+05
	DIS		.2299E+03	.5000E+03		
14	SOL	.138	.1054E+02	.6143E+03	.1389E+04	.2004E+04
	DIS		.6155E+02	.5000E+03		
15	SOL	.014	.3157E+03	.1839E+05	.4161E+05	.6000E+05
	DIS		.6155E+02	.5000E+03		
16	SOL	.012	.1052E+01	.6131E+02	.1387E+03	.2000E+03
	DIS		.6155E+02	.5000E+03		
20	SOL	10.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A1, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.1301E+06	.8776E+04	.1004E+05
PRIMARY.SETTLER.	.4364E+06	.2943E+05	.2469E+05
AERATIONTANKS...	.9391E+06	.6333E+05	.4519E+05
AIR.BLOWERS.....	.1367E+06	.9217E+04	.1783E+05
FINAL.SETTLER...	.4075E+06	.2748E+05	0
SLUDGE.PUMPS....	.5298E+05	.3573E+04	0
CONTROL.HOUSE...	.4208E+06	.2838E+05	0
SLUDGE.THICKENER	.1812E+06	.1222E+05	0
ANAEROB.DIGESTER	.3991E+06	.2692E+05	.1226E+05
ELUTRIATION.....	.1952E+06	.1316E+05	0
VACUUM..FILTER..	.3719E+06	.2508E+05	.4324E+05
SLUDGE.INCINERAT	.6796E+06	.4583E+05	.2099E+05
CHLORINATION....	.5562E+05	.3751E+04	.1945E+05
SITE.DEVELOPMENT	.6926E+05	.4671E+04	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.1882E+05	.5155E+00
PRIMARY.SETTLER.	.5412E+05	.1483E+01
AERATIONTANKS...	.1085E+06	.2973E+01
AIR.BLOWERS.....	.2705E+05	.7410E+00
FINAL.SETTLER...	.2748E+05	.7528E+00
SLUDGE.PUMPS....	.3573E+04	.9790E-01
CONTROL.HOUSE...	.2838E+05	.7776E+00
SLUDGE.THICKENER	.1222E+05	.3348E+00
ANAEROB.DIGESTER	.3917E+05	.1073E+01
ELUTRIATION.....	.1316E+05	.3606E+00
VACUUM..FILTER..	.6832E+05	.1872E+01
SLUDGE.INCINERAT	.6682E+05	.1831E+01
CHLORINATION....	.2320E+05	.6357E+00
SITE.DEVELOPMENT	.4671E+04	.1280E+00

TOTAL CAPITAL COST ($)	.4475E+07
TOTAL AMORTIZATION AND OPERATING COST($/YR)	.4955E+06
AMORTIZATION COST PER 1000 GAL (CENTS)	.8269E+01
OPERATING COST PER 1000 GAL (CENTS)	.5306E+01
TOTAL COST PER 1000 GAL (CENTS)	.1358E+02

Table A1, Continued

MLASS =	.7550E+03	MLBSS =	.2879E+03	MLNBSS =	.6226E+03
MLDSS =	.4691E+02	MLSS =	.2000E+04	VAER =	.2313E+01
VNIT =	.2944E+01	RETURN =	.4656E+00	MLISS =	.2844E+03
CWR =	.4066E+00	CARREM =	.7864E+00	BODREM =	.9018E+00
NITREM =	.1106E+00	PHOSREM =	.8573E-01	FRPS =	.5000E+00
URPS =	.4000E+03	URSS =	.3000E+01	XRSS =	.1242E-01
EFF =	.3798E+00	GPS =	.1375E+04	GSTH =	.9000E+01
ATHM =	.3335E+04	TRR =	.9500E+00	GE =	.8000E+03
GES =	.9000E+01	AE =	.1583E+04	ERR =	.7600E+00
WRE =	.3000E+01	TDIG =	.3300E+02	TD =	.1500E+02
VDIG =	.1526E+03	FRDIG =	.5910E+00	C1DIG =	.2605E+00
C2DIG =	.8550E+03	CH4CFD =	.9306E+05	CO2CFD =	.4887E+05
VFL =	.4900E+01	TVF =	.2380E+00	AVF =	.4418E+03
FOOD =	.1071E+03	DEGC =	.2000E+02	DO =	.1000E+01
VOLDIG =	0	ARATE =	.4500E-01	AYEARS =	.2500E+02
SOLIDS =	0	TSSLA =	0	DIGT =	0

```
CPTON =  .5202E+02
ECF   1.00 2.00 1.20 1.50 2.00 2.00 1.00
  1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG =  .6706E-01  CTRP =  .1000E+00  CTGO =  .1500E+00
CLAND =  .2000E-01  CCR =  .1337E+01
CCI =  .1570E+01       AF =  .6744E-01
```

TABLE A2

Computer Simulation of Activated Sludge Process, Deleting Primary Sedimentation with all other Variables Remaining Similar to Standard Activated Sludge

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISSOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	10.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	2000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

Table A2, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	10.478	.1136E+03	.1376E+03	.4007E+02	.1030E+02
	DIS		.4194E+02	.5786E+02	.1100E+02	.2052E+02
2	SOL	10.478	.1136E+03	.1376E+03	.4007E+02	.1030E+02
	DIS		.4194E+02	.5786E+02	.1100E+02	.2052E+02
5	SOL	10.093	.7183E+01	.6825E+01	.3763E+01	.1082E+01
	DIS		.1431E+02	.6187E+01	.1100E+02	.2542E+02
7	SOL	.386	.2110E+04	.1879E+04	.1106E+04	.1186E+03
	DIS		.1431E+02	.6187E+01	.1100E+02	.2542E+02
9	SOL	.478	.2946E+03	.8195E+02	.2507E+03	.1664E+02
	DIS		.1978E+02	.1642E+02	.1100E+02	.5235E+02
10	SOL	.386	.2110E+04	0	.1106E+04	.1186E+03
	DIS		.1431E+02	0	.1100E+02	.2542E+02
11	SOL	.354	.1150E+03	0	.6022E+02	.6463E+01
	DIS		.1431E+02	0	.1100E+02	.2542E+02
12	SOL	.032	.2443E+05	0	.1280E+05	.1373E+04
	DIS		.1431E+02	0	.1100E+02	.2542E+02
13	SOL	.032	.1300E+05	0	.1280E+05	.7362E+03
	DIS		.9842E+02	0	.1100E+02	.4396E+03
14	SOL	.112	.8893E+03	0	.8752E+03	.5034E+02
	DIS		.3534E+02	0	.1100E+02	.1290E+03
15	SOL	.015	.2100E+05	0	.2067E+05	.1189E+04
	DIS		.3534E+02	0	.1100E+02	.1290E+03
16	SOL	.013	.7000E+02	0	.6889E+02	.3963E+01
	DIS		.3534E+02	0	.1100E+02	.1290E+03
20	SOL	10.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

Table A2, Continued

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	10.478	.2044E+01	.3859E+02	.2421E+03	.2807E+03
	DIS		.4410E-01	.5000E+03		
2	SOL	10.478	.2044E+01	.3859E+02	.2421E+03	.2807E+03
	DIS		.4410E+01	.5000E+03		
5	SOL	10.093	.7183E-01	.3277E+01	.1365E+02	.1693E+02
	DIS		.5608E+01	.5000E+03		
7	SOL	.386	.2110E+02	.9628E+03	.4010E+04	.4973E+04
	DIS		.5608E+01	.5000E+03		
9	SOL	.478	.2960E+01	.2183E+03	.5891E+03	.8074E+03
	DIS		.1298E+02	.5000E+03		
10	SOL	.386	.2110E+02	.9628E+03	.4221E+04	.5184E+04
	DIS		.5608E+01	.5000E+03		
11	SOL	.354	.1150E+01	.5244E+02	.2299E+03	.2824E+03
	DIS		.5608E+01	.5000E+03		
12	SOL	.032	.2443E+03	.1114E+05	.4886E+05	.6000E+05
	DIS		.5608E+01	.5000E+03		
13	SOL	.032	.1309E+03	.1114E+05	.2601E+05	.3715E+05
	DIS		.1189E+03	.5000E+03		
14	SOL	.112	.8955E+01	.7621E+03	.1779E+04	.2541E+04
	DIS		.3394E+02	.5000E+03		
15	SOL	.015	.2115E+03	.1800E+05	.4200E+05	.6000E+05
	DIS		.3394E+02	.5000E+03		
16	SOL	.013	.7049E+00	.5999E+02	.1400E+03	.2000E+03
	DIS		.3394E+02	.5000E+03		
20	SOL	10.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A2, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.1301E+06	.8775E+04	.1004E+05
AERATIONTANKS...	.1448E+07	.9762E+05	.6457E+05
AIR.BLOWERS.....	.1900E+06	.1281E+05	.2613E+05
FINAL.SETTLER...	.4070E+06	.2745E+05	0
SLUDGE.PUMPS....	.5120E+05	.3453E+04	0
CONTROL.HOUSE...	.4208E+06	.2838E+05	0
SLUDGE.THICKENER	.1525E+06	.1029E+05	0
ANAEROB.DIGESTER	.3400E+06	.2293E+05	.1064E+05
ELUTRIATION.....	.2002E+06	.1350E+05	0
VACUUM..FILTER..	.3827E+06	.2581E+05	.4264E+05
SLUDGE.INCINERAT	.6922E+06	.4668E+05	.2165E+05
CHLORINATION....	.5562E+05	.3751E+04	.1945E+05
SITE.DEVELOPMENT	.6926E+05	.4671E+04	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.1881E+05	.5155E+00
AERATIONTANKS...	.1622E+06	.4443E+01
AIR.BLOWERS.....	.3895E+05	.1067E+01
FINAL.SETTLER...	.2745E+05	.7520E+00
SLUDGE.PUMPS....	.3453E+04	.9459E-01
CONTROL.HOUSE...	.2838E+05	.7775E+00
SLUDGE.THICKENER	.1029E+05	.2818E+00
ANAEROB.DIGESTER	.3357E+05	.9198E+00
ELUTRIATION.....	.1350E+05	.3698E+00
VACUUM..FILTER..	.6845E+05	.1875E+01
SLUDGE.INCINERAT	.6833E+05	.1872E+01
CHLORINATION....	.2321E+05	.6358E+00
SITE.DEVELOPMENT	.4671E+04	.1280E+00

TOTAL CAPITAL COST ($)	.4539E+07
TOTAL AMORTIZATION AND OPERATING COST ($/YR)	.5012E+06
AMORTIZATION COST PER 1000 GAL (CENTS)	.8387E+01
OPERATING COST PER 1000 GAL (CENTS)	.5346E+01
TOTAL COST PER 1000 GAL (CENTS)	.1373E+02

Table A2, Continued

MLASS =	.6534E+03	MLBSS =	.2760E+03	MLNBSS =	.7098E+03
MLDSS =	.3631E+02	MLSS =	.2000E+04	VAER =	.3676E+01
VNIT =	.5370E+01	RETURN =	.4399E+00	MLISS =	.3209E+03
CWR =	.5026E+00	CARREM =	.8670E+00	BODREM =	.9359E+00
NITREM =	.1720E+00	PHOSREM =	.1523E+00	FRPS =	.5000E+00
URPS =	.4000E+03	URSS =	.3000E+01	XRSS =	.1021E-01
EFF =	.3644E+00	GPS =	.1375E+04	GSTM =	.9000E+01
ATHM =	.2775E+04	TRR =	.9500E+00	GE =	.8000E+03
GES =	.9000E+01	AE =	.1633E+04	ERR =	.7600E+00
WRE =	.3000E+01	TDIG =	.3300E+02	TD =	.1500E+02
VDIG =	.1269E+03	FRDIG =	.4677E+00	C1DIG =	.2625E+00
C2DIG =	.8550E+03	CH4CFD =	.5881E+05	CO2CFD =	.3089E+05
VFL =	.4900E+01	TVF =	.2380E+00	AVF =	.4557E+03
FOOD =	.1575E+03	DEGC =	.2000E+02	DO =	.1000E+01
VOLDIG =	0	ARATE =	.4500E-01	AYEARS =	.2500E+02
SOLIDS =	0	TSSLA =	0	DIGT =	0

CPTON = .5053E+02
ECF 1.00 2.00 1.20 1.50 2.00 2.00 1.00
1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG = .6692E-01 CTRP = .1000E+00 CTGO = .1500E+00
CLAND = .2000E-01 CCR = .1337E+01
CCI = .1570E+01 AF = .6744E-01

TABLE A3

Simulation of Activated Sludge without Primary Sedimentation with Anaerobic Digestion and Vacuum Filtration

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	10.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	3000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	10.226	.1167E+03	.1411E+03	.4118E+02	.1054E+02
	DIS		.4264E+02	.5916E+02	.1100E+02	.2041E+02
2	SOL	10.226	.1167E+03	.1411E+03	.4118E+02	.1054E+02
	DIS		.4264E+02	.5916E+02	.1100E+02	.2041E+02
5	SOL	10.095	.6121E+01	.5879E+01	.3164E+01	.9097E+00
	DIS		.1481E+02	.7124E+01	.1100E+02	.2564E+02
7	SOL	.131	.6361E+04	.5745E+04	.3288E+04	.3436E+03
	DIS		.1481E+02	.7124E+01	.1100E+02	.2564E+02
9	SOL	.226	.6324E+03	.1800E+03	.5361E+03	.3434E+02
	DIS		.2659E+02	.2915E+02	.1100E+02	.8260E+02
10	SOL	.131	.6361E+04	0	.3288E+04	.3436E+03
	DIS		.1481E+02	0	.1100E+02	.2564E+02
11	SOL	.099	.4224E+03	0	.2184E+03	.2282E+02
	DIS		.1481E+02	0	.1100E+02	.2564E+02
12	SOL	.032	.2445E+05	0	.1264E+05	.1321E+04
	DIS		.1481E+02	0	.1100E+02	.2564E+02
13	SOL	.032	.1285E+05	0	.1264E+05	.6991E+03
	DIS		.9842E+02	0	.1100E+02	.4300E+03
14	SOL	.114	.8769E+03	0	.8628E+03	.4771E+02
	DIS		.3571E+02	0	.1100E+02	.1267E+03
15	SOL	.015	.2096E+05	0	.2062E+05	.1140E+04
	DIS		.3571E+02	0	.1100E+02	.1267E+03
16	SOL	.013	.6986E+02	0	.6873E+02	.3801E+01
	DIS		.3571E+02	0	.1100E+02	.1267E+03
20	SOL	10.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	10.226	.2096E+01	.3973E+02	.2485E+03	.2882E+03
	DIS		.4394E+01	.5000E+03		
2	SOL	10.226	.2096E+01	.3973E+02	.2485E+03	.2882E+03
	DIS		.4394E+01	.5000E+03		
5	SOL	10.095	.6121E-01	.2776E+01	.1163E+02	.1440E+02
	DIS		.5615E+01	.5000E+03		
7	SOL	.131	.6361E+02	.2884E+04	.1208E+05	.1497E+05
	DIS		.5615E+01	.5000E+03		
9	SOL	.226	.6355E+01	.4702E+03	.1265E+04	.1735E+04
	DIS		.2183E+02	.5000E+03		
10	SOL	.131	.6361E+02	.2884E+04	.1272E+05	.1561E+05
	DIS		.5615E+01	.5000E+03		
11	SOL	.099	.4224E+01	.1916E+03	.8448E+03	.1036E+04
	DIS		.5615E+01	.5000E+03		
12	SOL	.032	.2445E+03	.1109E+05	.4891E+05	.6000E+05
	DIS		.5615E+01	.5000E+03		
13	SOL	.032	.1294E+03	.1109E+05	.2570E+05	.3679E+05
	DIS		.1208E+03	.5000E+03		
14	SOL	.114	.8831E+01	.7568E+03	.1754E+04	.2511E+04
	DIS		.3440E+02	.5000E+03		
15	SOL	.015	.2110E+03	.1809E+05	.4191E+05	.6000E+05
	DIS		.3440E+02	.5000E+03		
16	SOL	.013	.7035E+00	.6029E+02	.1397E+03	.2000E+03
	DIS		.3440E+02	.5000E+03		
20	SOL	10.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A3, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.1302E+06	.8783E+04	.1004E+05
AERATIONTANKS...	.9931E+06	.6697E+05	.4728E+05
AIR.BLOWERS......	.1851E+06	.1248E+05	.2534E+05
FINAL.SETTLER...	.3286E+06	.2216E+05	0
SLUDGE.PUMPS....	.2541E+05	.1713E+04	0
CONTROL.HOUSE...	.4212E+06	.2840E+05	0
SLUDGE.THICKENER	.1561E+06	.1053E+05	0
ANAEROB.DIGESTOR	.3473E+06	.2342E+05	.1083E+05
ELUTRIATION.....	.2026E+06	.1366E+05	0
VACUUM..FILTER..	.3879E+06	.2616E+05	.4273E+05
SLUDGE.INCINERAT	.6986E+06	.4711E+05	.2194E+05
CHLORINATION....	.5567E+05	.3754E+04	.1945E+05
SITE.DEVELOPMENT	.6932E+05	.4675E+04	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.1882E+05	.5157E+00
AERATIONTANKS...	.1142E+06	.3130E+01
AIR.BLOWERS......	.3782E+05	.1036E+01
FINAL.SETTLER...	.2216E+05	.6071E+00
SLUDGE.PUMPS....	.1713E+04	.4694E-01
CONTROL.HOUSE...	.2840E+05	.7782E+00
SLUDGE.THICKENER	.1053E+05	.2884E+00
ANAEROB.DIGESTOR	.3425E+05	.9384E+00
ELUTRIATION.....	.1366E+05	.3743E+00
VACUUM..FILTER..	.6889E+05	.1887E+01
SLUDGE.INCINERAT	.6905E+05	.1892E+01
CHLORINATION....	.2321E+05	.6359E+00
SITE.DEVELOPMENT	.4675E+04	.1281E+00

TOTAL CAPITAL COST	.4001E+07
TOTAL AMORTIZATION AND OPERATING COST	.4474E+06
AMORTIZATION COST PER 1000 GAL	.7392E+01
OPERATING COST PER 1000 GAL	.4866E+01
TOTAL COST PER 1000 GAL	.1226E+02

Table A3, Continued

MLASS =	.9463E+03	MLBSS =	.4625E+03	MLNBSS =	.1062E+04
MLDSS =	.5237E+02	MLSS =	.3000E+04	VAER =	.2455E+01
VNIT =	.3433E+01	RETURN =	.1835E+00	MLISS =	.4807E+03
CWR =	.5132E+00	CARREM =	.8703E+00	BODREM =	.9359E+00
NITREM =	.1530E+00	PHOSREM =	.1367E+00	FRPS =	.5000E+00
URPS =	.4000E+03	URSS =	.6000E+01	XRSS =	.5774E-02
EFF =	.3571E+00	GPS =	.1375E+04	GSTH =	.9000E+01
ATHM =	.2841E+04	TRR =	.9500E+00	GE =	.8000E+03
GES =	.9000E+01	AE =	.1655E+04	ERR =	.7600E+00
WRE =	.3000E+01	TDIG =	.3300E+02	TD =	.1500E+02
VDIG =	.1299E+03	FRDIG =	.4745E+00	C1DIG =	.2605E+00
C2DIG =	.8550E+03	CH4CFD =	.6116E+05	CO2CFD =	.3212E+05
VFL =	.4900E+01	TVF =	.2380E+00	AVF =	.4620E+03
FOOD =	.1566E+03	DEGC =	.2000E+02	DO =	.1000E+01
VOLDIG =	0	ARATE =	.4500E-01	AYEARS =	.2500E+02
SOLIDS =	0	TSSLA =	0	DIGT =	0

CPTON = .5015E+02
ECF 1.00 2.00 1.20 1.50 2.00 2.00 1.00
1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG = .6716E-01 CTRP = .1000E+00 CTGO = .1500E+00
CLAND = .2000E-01 CCR = .1337E+01
CCI = .1570E+01 AF = .6744E-01

TABLE A4

Computer Simulation of Activated Sludge with Aerobic Digestion, 60% Volatile Solids Destruction and 6% Suspended Solids Held in the Digester

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISSOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	10.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	2000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

Table A4, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	10.174	.1067E+03	.1384E+03	.3273E+02	.1013E+02
	DIS		.4252E+02	.5894E+02	.1100E+02	.2358E+02
2	SOL	10.161	.5343E+02	.6927E+02	.1638E+02	.5071E+01
	DIS		.4252E+02	.5894E+02	.1100E+02	.2358E+02
5	SOL	9.998	.9148E+01	.9967E+01	.4186E+01	.1494E+01
	DIS		.1262E+02	.3032E+01	.1100E+02	.2468E+02
7	SOL	.164	.2197E+04	.2228E+04	.1005E+04	.1546E+03
	DIS		.1262E+02	.3032E+01	.1100E+02	.2468E+02
8	SOL	.013	.4269E+05	.5535E+05	.1309E+05	.4051E+04
	DIS		.4252E+02	.5894E+02	.1100E+02	.2358E+02
9	SOL	.174	.2058E+03	.3079E+02	.1894E+03	.1752E+02
	DIS		.1477E+02	.7059E+01	.1100E+02	.2864E+03
10	SOL	.177	.5114E+04	0	.1876E+04	.4353E+03
	DIS		.1477E+02	0	.1100E+02	.2460E+02
11	SOL	.160	.2176E+03	0	.2002E+03	.1853E+02
	DIS		.1477E+02	0	.1100E+02	.2864E+03
12	SOL	.177	.5114E+04	0	.1876E+04	.4353E+03
	DIS		.1477E+02	0	.1100E+02	.2460E+02
13	SOL	.017	.2176E+05	0	.2002E+05	.1834E+04
	DIS		.1477E+02	0	.1100E+02	.2864E+03
15	SOL	.017	.2176E+05	0	.2002E+05	.1834E+04
	DIS		.1477E+02	0	.1100E+02	.2864E+03
16	SOL	.014	.7254E+02	0	.6674E+02	.6114E+01
	DIS		.1477E+02	0	.1100E+02	.2864E+03
20	SOL	10.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

Table A4, Continued

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	10.174	.2020E+01	.2980E+02	.2273E+03	.2571E+03
	DIS		.4832E+01	.5000E+03		
2	SOL	10.161	.1011E+01	.1492E+02	.1138E+03	.1287E+03
	DIS		.4832E+01	.5000E+03		
5	SOL	9.998	.9148E-01	.3073E+01	.1738E+02	.2045E+02
	DIS		.5399E+01	.5000E+03		
7	SOL	.164	.2197E+02	.7380E+03	.4174E+04	.4912E+04
	DIS		.5399E+01	.5000E+03		
8	SOL	.013	.8080E+03	.1192E+05	.9093E+05	.1029E+06
	DIS		.4832E+01	.5000E+03		
9	SOL	.174	.3162E+01	.1861E+02	.4116E+03	.4303E+03
	DIS		.5261E+02	.5000E+03		
10	SOL	.177	.7859E+02	.1544E+04	.1023E+05	.1177E+05
	DIS		.5358E+01	.5000E+03		
11	SOL	.160	.3344E+01	.1544E+02	.4352E+03	.6000E+03
	DIS		.5261E+02	.5000E+03		
12	SOL	.177	.7859E+02	.1544E+04	.1023E+05	.1177E+05
	DIS		.5358E+01	.5000E+03		
13	SOL	.017	.3311E+03	.1633E+05	.4352E+05	.6000E+05
	DIS		.5261E+02	.5000E+03		
15	SOL	.017	.3311E+03	.1633E+05	.4352E+05	.6000E+05
	DIS		.5261E+02	.5000E+03		
16	SOL	.014	.1104E+01	.5442E+02	.1451E+03	.2000E+03
	DIS		.5261E+02	.5000E+03		
20	SOL	10.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A4, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.1302E+06	.8780E+04	.1004E+05
PRIMARY.SETTLER.	.4307E+06	.2904E+05	.2441E+05
AERATIONTANKS...	.9163E+06	.6179E+05	.5685E+05
AIR.BLOWERS.....	.1833E+06	.1236E+05	.2508E+05
FINAL.SETTLER...	.4032E+06	.2719E+05	0
SLUDGE.PUMPS....	.5277E+05	.3559E+04	0
CONTROL.HOUSE...	.4210E+06	.2839E+05	0
AEROBIC.DIGESTER	.3950E+06	.2664E+05	0
VACUUM..FILTER..	.4223E+06	.2848E+05	.4800E+05
SLUDGE.INCINERAT	.7376E+06	.4974E+05	.2402E+05
CHLORINATION....	.5565E+05	.3753E+04	.1945E+05
SITE.DEVELOPMENT	.6929E+05	.4673E+04	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.1882E+05	.5156E+00
PRIMARY.SETTLER.	.5346E+05	.1465E+01
AERATIONTANKS...	.1186E+06	.3250E+01
AIR.BLOWERS.....	.3744E+05	.1026E+01
FINAL.SETTLER...	.2719E+05	.7449E+00
SLUDGE.PUMPS....	.3559E+04	.9750E-01
CONTROL.HOUSE...	.2839E+05	.7779E+00
AEROBIC.DIGESTER	.2664E+05	.7298E+00
VACUUM..FILTER..	.7648E+05	.2095E+01
SLUDGE.INCINERAT	.7376E+05	.2021E+01
CHLORINATION....	.2321E+05	.6358E+00
SITE.DEVELOPMENT	.4673E+04	.1280E+00

TOTAL CAPITAL COST ($)	.4217E+07
TOTAL AMORTIZATION AND OPERATING COST($/YR)	.4923E+06
AMORTIZATION COST PER 1000 GAL(CENTS)	.7792E+01
OPERATING COST PER 1000 GAL (CENTS	.5695E+01
TOTAL COST PER 1000 GAL (CENTS)	.1349E+02

Table A4, Continued

MLASS =	.8516E+03	MLBSS =	.2681E+03	MLNBSS =	.5753E+03
MLDSS =	.5932E+02	MLSS =	.2000E+04	VAER =	.2251E+01
VNIT =	.2514E+01	RETURN =	.4697E+00	MLISS =	.2460E+03
CWR =	.3713E+00	CARREM =	.7768E+00	BODREM =	.9003E+00
NITREM =	.1009E+00	PHOSREM =	.7551E-01	FRPS =	.5000E+00
URPS =	.4000E+03	URSS =	.3000E+01	XRSS =	.1249E-01
EFF =	.4055E+00	GPS =	.1375E+04	GSTH =	.9000E+01
ATHM =	0	TRR =	.9500E+00	GE =	.8000E+03
GES =	.9000E+01	AE =	0	ERR =	.7600E+00
WRE =	.3000E+01	TDIG =	.3300E+02	TD =	.1500E+02
VDIG =	0	FRDIG =	0	C1DIG =	0
C2DIG =	0	CH4CFD =	0	CO2CFD =	0
VFL =	.4900E+01	TVF =	.2380E+00	AVF =	.5061E+03
FOOD =	.1096E+03	DEGC =	.2000E+02	DO =	.1000E+01
VOLDIG =	.8713E+00	ARATE =	.4500E-01	AYEARS =	.2500E+02
SOLIDS =	.6000E+01	TSSLA =	.1000E-01	DIGT =	.4300E+02

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CPTON =   .5083E+02
ECF    1.00 2.00 1.20 1.50 2.00 2.00 1.00
  1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG =   .6765E-01   CTRP =   .1000E+00   CTGO =   .1500E+00
CLAND =   .2000E-01   CCR =   .1338E+01
CCI =   .1570E+01       AF =   .6744E-01
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TABLE A5

Standard Activated Sludge Process with Sand Drying Beds

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISSOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	10.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	2000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

Table A5, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	10.173	.1077E+03	.1433E+03	.3105E+02	.1021E+02
	DIS		.4248E+02	.5892E+02	.1097E+02	.1896E+02
2	SOL	10.160	.5390E+02	.7172E+02	.1555E+02	.5112E+01
	DIS		.4248E+02	.5892E+02	.1097E+02	.1896E+02
5	SOL	9.996	.8993E+01	.9986E+01	.4024E+01	.1485E+01
	DIS		.1257E+02	.2990E+01	.1097E+02	.2007E+02
7	SOL	.164	.2180E+04	.2252E+04	.9753E+03	.1577E+03
	DIS		.1257E+02	.2990E+01	.1097E+02	.2007E+02
8	SOL	.013	.4306E+05	.5730E+05	.1242E+05	.4085E+04
	DIS		.4248E+02	.5892E+02	.1097E+02	.1896E+02
9	SOL	.173	.2614E+03	.3172E+03	.9182E+02	.2248E+02
	DIS		.1220E+02	.5818E+01	.9093E+01	.1657E+02
10	SOL	.177	.5124E+04	0	.1800E+04	.4405E+03
	DIS		.1472E+02	0	.1097E+02	.1999E+02
11	SOL	.143	.3153E+03	0	.1108E+03	.2711E+02
	DIS		.1472E+02	0	.1097E+02	.1999E+02
12	SOL	.033	.2595E+05	0	.9113E+04	.2231E+04
	DIS		.1472E+02	0	.1097E+02	.1999E+02
13	SOL	.033	.9320E+04	0	.9113E+04	.8092E+03
	DIS		.9839E+02	0	.1097E+02	.9440E+03
20	SOL	10.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

Table A5, Continued

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	10.173	.2034E+01	.3088E+02	.2293E+03	.2602E+03
	DIS		.3996E+01	.4985E+03		
2	SOL	10.160	.1018E+01	.1546E+02	.1148E+03	.1303E+03
	DIS		.3996E+01	.4985E+03		
5	SOL	9.996	.8993E-01	.3160E+01	.1709E+02	.2025E+02
	DIS		.4574E+01	.4985E+03		
7	SOL	.164	.2180E+02	.7658E+03	.4141E+04	.4907E+04
	DIS		.4574E+01	.4985E+03		
8	SOL	.013	.8138E+03	.1235E+05	.9173E+05	.1041E+06
	DIS		.3996E+01	.4985E+03		
9	SOL	.173	.4022E+01	.8164E+02	.5229E+03	.6045E+03
	DIS		.3758E+01	.4133E+03		
10	SOL	.177	.7883E+02	.1600E+04	.1025E+05	.1185E+05
	DIS		.4533E+01	.4985E+03		
11	SOL	.143	.4852E+01	.9848E+02	.6307E+03	.7292E+03
	DIS		.4533E+01	.4985E+03		
12	SOL	.033	.3992E+03	.8103E+04	.5190E+05	.6000E+05
	DIS		.4533E+01	.4985E+03		
13	SOL	.033	.1448E+03	.8103E+04	.1864E+05	.2674E+05
	DIS		.2589E+03	.4985E+03		
20	SOL	10.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A5, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.1303E+06	.8788E+04	.1004E+05
PRIMARY.SETTLER.	.4311E+06	.2907E+05	.2441E+05
AERATIONTANKS...	.9349E+06	.6305E+05	.4497E+05
AIR.BLOWERS.....	.1439E+06	.9703E+04	.1892E+05
FINAL.SETTLER...	.4035E+06	.2721E+05	0
SLUDGE.PUMPS....	.5282E+05	.3562E+04	0
CONTROL.HOUSE...	.4214E+06	.2842E+05	0
SLUDGE.THICKENER	.1595E+06	.1075E+05	0
ANAEROB.DIGESTER	.3542E+06	.2389E+05	.1102E+05
SLUDGE.INCINERAT	0	0	0
SLUDGE.DRYING.BD	.5273E+06	.3556E+05	.8422E+04
CHLORINATION....	.5570E+05	.3756E+04	.1945E+05
SITE.DEVELOPMENT	.6936E+05	.4678E+04	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.1883E+05	.5159E+00
PRIMARY.SETTLER.	.5348E+05	.1465E+01
AERATIONTANKS...	.1080E+06	.2959E+01
AIR.BLOWERS.....	.2862E+05	.7842E+00
FINAL.SETTLER...	.2721E+05	.7456E+00
SLUDGE.PUMPS....	.3562E+04	.9760E-01
CONTROL.HOUSE...	.2842E+05	.7787E+00
SLUDGE.THICKENER	.1075E+05	.2946E+00
ANAEROB.DIGESTER	.3491E+05	.9563E+00
SLUDGE.INCINERAT	0	0
SLUDGE.DRYING.BD	.4399E+05	.1205E+01
CHLORINATION....	.2321E+05	.6358E+00
SITE.DEVELOPMENT	.4678E+04	.1282E+00

TOTAL CAPITAL COST ($)	.3684E+07
TOTAL AMORTIZATION AND OPERATING COST ($)	.3857E+06
AMORTIZATION COST PER 1000 GAL (CENTS)	.6807E+01
OPERATING COST PER 1000 GAL (CENTS)	.3760E+01
TOTAL COST PER 1000 GAL (CENTS)	.1057E+02

Table A5, Continued

```
 MLASS =  .8684E+03    MLBSS =  .2649E+03   MLNBSS =  .5467E+03
 MLDSS =  .6188E+02     MLSS =  .2000E+04     VAER =  .2299E+01
  VNIT =  .2644E+01   RETURN =  .4697E+00    MLISS =  .2553E+03
   CWR =  .3668E+00   CARREM =  .7799E+00   BODREM =  .9023E+00

 NITREM =  .1191E+00  PHOSREM =  .8481E-01    FRPS =  .5000E+00
  URPS =  .4000E+03     URSS =  .3000E+01     XRSS =  .1238E-01
   EFF =  .4131E+00      GPS =  .1375E+04     GSTH =  .9000E+01
  ATHM =  .2904E+04      TRR =  .9500E+00       GE =  .8000E+03

   GES =  .9000E+01       AE =      0          ERR =  .7600E+00
   WRE =  .3000E+01     TDIG =  .3300E+02       TD =  .1500E+02
  VDIG =  .1329E+03    FRDIG =  .6408E+00    C1DIG =  .2605E+00
 C2DIG =  .8550E+03   CH4CFD =  .8979E+05   CO2CFD =  .4716E+05

  VFL =  .4900E+01      TVF =  .2380E+00      AVF =      0
  FOOD =  .1123E+03     DEGC =  .2000E+02       DO =  .1000E+01

VOLDIG =      0        ARATE =  .4500E-01   AYEARS =  .2500E+02
SOLIDS =      0        TSSLA =      0         DIGT =      0

 CPTON =      0
ECF    1.00 2.00 1.20 1.50 2.00 2.00 1.00
  1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG =  .6925E-01  CTRP =  .1000E+00  CTGO =  .1500E+00
CLAND =  .2000E-01  CCR =  .1339E+01
CCI =  .1570E+01      AF =  .6744E-01
```

TABLE A6

Simulation of the Lee, Massachusetts Treatment Plant

INPUT PARAMETERS

SOLID BOD (MG/L)	140.25
DISSOLVED BOD (MG/L)	59.84
TOTAL SUSPENDED SOLIDS (MG/L)	253.65
VOLATILE SUSPENDED SOLIDS (MG/L)	223.65
SEWAGE VOLUME (MGD)	1.00
MIXED LIQUOR SUSPENDED SOLIDS HELD IN AERATOR (MG/L)	3000.00
MAXIMUM BOD IN EFFLUENT (MG/L)	13.00

Table A6, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STATION		MGD	CARBON	BOD	NBIOCARBON	NITROGEN
1	SOL	1.023	.1072E+03	.1351E+03	.3494E+02	.1004E+02
	DIS		.4237E+02	.5866E+02	.1100E+02	.2164E+02
2	SOL	1.023	.1072E+03	.1351E+03	.3494E+02	.1004E+02
	DIS		.4237E+02	.5866E+02	.1100E+02	.2164E+02
5	SOL	1.000	.6380E+01	.6399E+01	.3165E+01	.9656E+00
	DIS		.1454E+02	.6617E+01	.1100E+02	.2656E+02
7	SOL	.023	.3264E+04	.3076E+04	.1619E+04	.1862E+03
	DIS		.1454E+02	.6617E+01	.1100E+02	.2656E+02
9	SOL	.023	.2040E+03	-.9314E+02	.2538E+03	.1163E+02
	DIS		.1454E+02	.6617E+01	.1100E+02	.1385E+03
10	SOL	.023	.3264E+04	0	.1619E+04	.1862E+03
	DIS		.1454E+02	0	.1100E+02	.2656E+02
11	SOL	.021	.2040E+03	0	.2538E+03	.1164E+02
	DIS		.1454E+02	0	.1100E+02	.1385E+03
12	SOL	.023	.3264E+04	0	.1619E+04	.1862E+03
	DIS		.1454E+02	0	.1100E+02	.2656E+02
13	SOL	.001	.2040E+05	0	.2538E+05	.1152E+04
	DIS		.1454E+02	0	.1100E+02	.1385E+03
15	SOL	.001	.2040E+05	0	.2538E+05	.1152E+04
	DIS		.1454E+02	0	.1100E+02	.1385E+03
16	SOL	.001	.2040E+03	0	.2538E+03	.1152E+02
	DIS		.1454E+02	0	.1100E+02	.1385E+03
20	SOL	1.000	.1050E+03	.1403E+03	.3000E+02	.1000E+02
	DIS		.4300E+02	.5984E+02	.1100E+02	.1900E+02

Table A6, Continued

STATION		MGD	PHOSPHORUS	FIXED MATTER	VSS	TSS
1	SOL	1.023	.2001E+01	.2980E+02	.2283E+03	.2581E+03
	DIS		.4471E+01	.5000E+03		
2	SOL	1.023	.2001E+01	.2980E+02	.2283E+03	.2581E+03
	DIS		.4471E+01	.5000E+03		
5	SOL	1.000	.6380E-01	.2393E+01	.1212E+02	.1452E+02
	DIS		.5677E+01	.5000E+03		
7	SOL	.023	.3264E+02	.1224E+04	.6202E+04	.7426E+04
	DIS		.5677E+01	.5000E+03		
9	SOL	.023	.2039E+01	.2100E+02	.4081E+03	.4291E+03
	DIS		.2530E+02	.5000E+03		
10	SOL	.023	.3264E+02	.1224E+04	.6528E+04	.7752E+04
	DIS		.5677E+01	.5000E+03		
11	SOL	.021	.2040E+01	.1224E+02	.4081E+03	.6000E+03
	DIS		.2530E+02	.5000E+03		
12	SOL	.023	.3264E+02	.1224E+04	.6528E+04	.7752E+04
	DIS		.5677E+01	.5000E+03		
13	SOL	.001	.2020E+03	.1901E+05	.4081E+05	.6000E+05
	DIS		.2530E+02	.5000E+03		
15	SOL	.001	.2020E+03	.1901E+05	.4081E+05	.6000E+05
	DIS		.2530E+02	.5000E+03		
16	SOL	.001	.2020E+01	.1901E+03	.4081E+03	.6000E+03
	DIS		.2530E+02	.5000E+03		
20	SOL	1.000	.2000E+01	.3000E+02	.2236E+03	.2536E+03
	DIS		.4000E+01	.5000E+03		

Table A6, Continued

CASE NO. 1 ACTIVATED SLUDGE PROCESS CALCULATION JUNE 1969

STRUCTURE	CAP COST DOLLARS	AMORT. $/YEAR	OPERATION $/YEAR
PRELIMINARY.TRT.	.2559E+05	.1725E+04	.2650E+04
AERATIONTANKS...	.1211E+06	.8165E+04	.1143E+05
AIR.BLOWERS.....	.3526E+05	.2378E+04	.3167E+04
FINAL.SETTLER...	.4167E+05	.2810E+04	0
SLUDGE.PUMPS....	.1007E+05	.6788E+03	0
CONTROL.HOUSE...	.6962E+05	.4695E+04	0
AEROBIC.DIGESTER	.6622E+05	.4466E+04	0
SLUDGE.INCINERAT	0	0	0
SLUDGE.DRYING.BD	.2444E+05	.1648E+04	.7487E+03
CHLORINATION....	.1566E+05	.1056E+04	.1945E+04
SITE.DEVELOPMENT	.7658E+04	.5165E+03	0

STRUCTURE	AMORTIZATION PLUS OPERATING COST $/YEAR	CENTS/1000 GAL
PRELIMINARY.TRT.	.4375E+04	.1199E+01
AERATIONTANKS...	.1960E+05	.5369E+01
AIR.BLOWERS.....	.5544E+04	.1519E+01
FINAL.SETTLER...	.2810E+04	.7699E+00
SLUDGE.PUMPS....	.6788E+03	.1860E+00
CONTROL.HOUSE...	.4695E+04	.1286E+01
AEROBIC.DIGESTER	.4466E+04	.1223E+01
SLUDGE.INCINERAT	0	0
SLUDGE.DRYING.BD	.2397E+04	.6567E+00
CHLORINATION....	.3001E+04	.8223E+00
SITE.DEVELOPMENT	.5165E+03	.1415E+00

TOTAL CAPITAL COST	.4172E+06
TOTAL AMORTIZATION AND OPERATING COST	.4808E+05
AMORTIZATION COST PER 1000 GAL	.7709E+01
OPERATING COST PER 1000 GAL	.5464E+01
TOTAL COST PER 1000 GAL	.1317E+02

Table A6, Continued

MLASS =	.1025E+04	MLBSS =	.4860E+03	MLNBSS =	.1019E+04
MLDSS =	.6200E+02	MLSS =	.3000E+04	VAER =	.2370E+00
VNIT =	.3114E+00	RETURN =	.4635E+00	MLISS =	.4080E+03
CWR =	.4918E+00	CARREM =	.8633E+00	BODREM =	.9343E+00
NITREM =	.1507E+00	PHOSREM =	.1328E+00	FRPS =	.5000E+00
URPS =	.4000E+03	URSS =	.3000E+01	XRSS =	.5864E-02
EFF =	.3715E+00	GPS =	.1375E+04	GSTH =	.9000E+01
ATHM =	0	TRR =	.9500E+00	GE =	.8000E+03
GES =	.9000E+01	AE =	0	ERR =	.7600E+00
WRE =	.3000E+01	TDIG =	.3300E+02	TD =	.1500E+02
VDIG =	.5853E+01	FRDIG =	0	C1DIG =	0
C2DIG =	0	CH4CFD =	0	CO2CFD =	0
VFL =	.4900E+01	TVF =	.2380E+00	AVF =	0
FOOD =	.1532E+03	DEGC =	.2000E+02	DO =	.1000E+01
VOLDIG =	.8440E-01	ARATE =	.4500E-01	AYEARS =	.2500E+02
SOLIDS =	.6000E+01	TSSLA =	.1000E-01	DIGT =	.4300E+02

CPTON = 0
ECF 1.00 2.00 1.20 1.50 2.00 2.00 1.00
1.50 2.00 1.50 1.00 1.00 1.00 1.00 1.00
CENG = .9511E-01 CTRP = .1000E+00 CTCO = .1500E+00
CLAND = .2000E-01 CCR = .1365E+01
CCI = .1275E+01 AF = .6744E-01

APPENDIX II

Computer Program Variable Definitions

APPENDIX II

Computer Variable Listing and Definition:

VARIABLE NAME	DEFINITION
ACOST (i)	Amortization cost of process i, \$/yr.
AE	Elutriation tank surface area, sq ft
AEFF	Efficiency of aerator diffusers, corrected for water temperature and dissolved oxygen deficit
AEFF20	Efficiency of aerator diffusers at zero dissolved oxygen and 20^{o}C
AF	Amortization factor, capital cost recovery factor at interest of ARATE for AYEARS
AFS	Final settler surface area, sq ft/1000
AIRCFD	Aerator air requirement in standard cubic feet per day
ALK(i)	Alkalinity as $CaCO_3$ at station i, mg/l
AOCOST(i)	Amortization plus operating costs for process i, \$/yr
APS	Primary settler surface area, sq ft/1000
ARATE	Interest rate at which the plant is amortized (%)
AREA	Area required for sludge drying beds, acres
ASB	Area of the sand drying beds, as in the original Smith model, sq ft
ASMAX	Current maximum value of XMLASS, mg/l
ASMIN	Current minimum value of XMLASS, mg/l
ASS(i)	The ith input value of XMLSS
ATH1	Surface area of the thickener, based upon overflow rate, sq ft

APPENDIX II, continued

VARIABLE NAME	DEFINITION
ATH2	Surface area of the thickener based on solids loading, sq ft
ATHM	The larger value of ATH1 and ATH2
AVF	Area of Vacuum filter, sq ft
AYEARS	Number of years for amortization
BOD(i)	Total BOD at station i, mg/l oxygen
BODREM	Fraction of 5-day BOD removed by the aerator
BSIZE	Required blower size, cfm.
CAER20	Rate constant for BOD removal in the aerator at 20^{o}C.
CAER	Rate constant for BOD removal corrected for water temperature
CAIRP	Cost of electricity for air blowers, $/yr.
CAPKG	Capital cost per thousand gallons treated, cents
C1DIG	Rate constant for anaerobic digester
C2DIG	Rate constant for anaerobic digester
CCI	Capital cost index, ratio of ENR Index of present to 812
CCOST(i)	Capital cost of ith process, $
CCR	Capital cost ratio: 1 + CENG + CTRP + CLAND + CTGO

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
CEDR	Rate of solids destruction in the aerator, fraction/day
CENG	Cost of engineering, fraction of total cost
CFECL3	Cost of ferric chloride, $/lb
CFPGAL	Aerator air requirement, in standard cubic feet/gallon raw sewage
CH4CFD	Standard cubic feet of methane produced each day by anaerobic digester
CKWH	Cost of electricity, $/kilowatt hour
CNIT	Rate constant for nitrification in aerator
CLAND	Cost of land expressed as fraction of capital cost
CO2CFD	Standard cubic feet of carbon dioxide produced daily in anaerobic digester
COPKG	Operating cost per 1000 gallons, cents
COSTO(i)	Operating cost of ith process, $/yr
CPERKG(i)	Total cost of ith process, cents/1000 gallons
CPTON	Cost of sludge dewatering per ton of dry solids
CREM	Fraction of organic carbon removed by the aerator
CTGO	Contingincy cost, fraction of total capital cost
CTRP	Contractor's profit, fraction of total capital cost

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
CWR	Ratio of carbon in final settler sludge to carbon at influent
DBOD(i)	Dissolved BOD at station i, mg/l
DEGC	Water temperature, degrees centigrade
DEMBOD(i)	The ith input value for the required maximum BOD at station 5, mg/l
DFM(i)	Concentration of dissolved fixed matter at station i
DIGT	Digestion time for aerobic digester, days
DIGC12	Concentration of biodegradable carbon at station 12, mg/l
DIGC13	Concentration of biodegradable carbon at station 13,mg/l
DN(i)	Dissolved nitrogen at station i, mg/l
DNBC(i)	Dissolved non-biodegradable carbon at station i, mg/l
DO	Concentration of dissolved oxygen in the aerator, mg/l
DOC(i)	Dissolved organic carbon at station i, mg/l
DOSAT	Oxygen saturation level at the mid-depth of aerator, mg/l
DP(i)	Dissolved phosphorus level at station i, mg/l
DRYDAY	Number of days suitable for sludge drying on sand beds, days
ECF(i)	Excess capacity factor for process i, ratio of required capacity to design capacity

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
EFF	Efficiency of the aerator: pounds of carbon escaping to the atmosphere per pound of carbon entering the aerator
ERR	Solids recovery ratio for the elutriation tank. Total solids in stream 15 divided by total solids in stream 13
ERROR	Difference between FMAX and FMIN
FECL3	Concentration of ferric chloride used for sludge conditioning, mg/l
FMAX	Current maximum value of FOOD, mg/l 5-day BOD
FMIN	Current minimum value of FOOD, mg/l 5-day BOD
FOOD	5-day BOD synthesized into active solids in the aerator per day, mg/l oxygen
FRDIG	Fraction of carbon entering anaerobic digester which is converted to gas
FRPS	Fraction of solids removed in primary settler
GE	Design overflow of elutriation tank gallons per day per sq ft
GES	Design solids loading of elutriation tank, lb/day/sq ft
GPS	Primary settler overflow rate, gpd/sq ft
GSS	Design overflow rate for secondary settler, gpd/sq ft
GSTH	Thickener solids loading rate, lb/day/sq ft
GTH	Design overflow rate for thickener, gpd/sq ft
HC	Reference head corresponding to Rc,cm

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
HO	Depth of sludge filling plus sand depth on sand drying beds, cm.
LOOPS	Number of iterations carried out when stream number 9 is returned
NAS	Number of input values for MLSS
NBOD	Number of input values for 5-day BOD
NFORK(i)	Decision matrix which decides which flow pattern is to be used
NOTONS	Number of tons of dry solids applied on each run of the sand drying beds
OUTPUT(i)	Decision matrix which decides which portion of the simulation is to be printed
PREPT	Time required to clean and prepare sand beds between applications, days
Q(i)	Flow at station i, MGD
RC	Specific resistance of the sludge, sec^2/gm
REMN	Fraction of nitrogen removed from the main stream by the aerator
REMP	Fraction of phosphorus removed from the main stream by the aerator
RETURN	Sludge return ratio, Q_6/Q_2
SAND	Depth of the sand drying bed, cm
SBOD(i)	Solid 5-day BOD at station i, mg/l
SCI	Constant rate drying constant, kg/m^2-hr
SFM(i)	Solid fixed matter at station i, mg/l
SIGMA	Coefficient of compressibility of the sludge

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
SNBC(i)	Solid non-biodegradable carbon at station i, mg/l
SO	Original solids content of the sludge, decimal fraction
SOC(i)	Solid carbon concentration at station i, mg/l
SON(i)	Solid nitrogen at station i, mg/l
Solids	Total suspended solids level maintained in the aerobic digester, mg/l
SOP(i)	Solid phosphorus at station i, mg/l
STF	Fractional solids content of sludge after drying on drying beds
STO	Fractional solids content of sludge after draining, when drying begins on sand drying beds
T	Time required for sludge to drain on sand drying beds, hr
TA	Aerator detention time, VAER/Q_2, days
TAN	Aerator detention time required to achieve nitrification, days
TACOST	Total amortization cost, $/yr
TDRY	Time required for sludge to dry on sand drying beds, hr
TBOD2	Total 5-day BOD at station 2, mg/l
TBOD5	Total 5-day BOD at station 5, mg/l
TCOST	Total treatment cost, ¢/1000 gal
TCOSTO	Total operating and maintenance cost for the plant, $/yr
TD	Anaerobic digester detention time, days

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
TDIG	Sludge temperature in the anaerobic digester, degrees centigrade
TEMP1	Temporary storage space
TOC2	Total carbon concentration at station 2, mg/l
TOC5	Total carbon concentration at station 5, mg/l
TOC7	Total carbon concentration at station 7, mg/l
TOL	Tolerance allowed between calculated XMLSS and the required value
TOTAOC	Total amortization and operating cost per year, $/yr
TOTCC	Total capital cost of plant, $
TRR	Solids recovery ratio for the thickener, TSS(12/TSS(10)
TSS(i)	Total suspended solids at station i, mg/l
TSSLA	Fraction of total suspended solids lost in the aerobic digester supernatant
TVF	Fraction of the time the vacuum filters are in operation
VOLUME	Number of gallons per day of sludge fed to the aerobic digester
UC	Moisture content of the sludge on the drying bed when the falling rate period is begun
UO	Moisture content of sludge on the drying bed at the time of application
URPS	Ratio of solids in stream 8 to solids in stream 1, TSS(8)/TSS(1)

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
URSS	Ratio of solids in stream 7 to the solids in aerator, TSS(7)/MLSS
VAER	Volume of aerator, millions of gallons
VDIG	Volume of anaerobic digester, thousands of cubic feet
VFL	Vacuum filter loading, gal/hr/sq ft
VNIT	Volume of aerator required to achieve nitrification, millions of gallons
VOLDIG	Volume of aerobic digester, millions of gallons
VOLRED	Per cent reduction of influent volatile solids in the aerobic digester
VSS(i)	Concentration of volatile suspended solids at station i, mg/l
WAS	Pounds of active solids in aerator
WDN	Pounds per day of dissolved nitrogen in streams 5 and 7 from the aerator, lb/day
WFOOD	Pounds per day of 5-day BOD synthesized to active solids in the aerator
WP	Per cent moisture in filtered sludge
WRE	Wash water ratio for elutriation, Q(17)/Q(13)
WTS	Weight of total solids per unit area, gm/sq m
XLANDC	Cost of land for the sand drying beds, $/acre

APPENDIX II, Continued

VARIABLE NAME	DEFINITION
XLNBSS	Concentration of inert organic solids in aerator caused by inert organic solids in the influent, mg/l
XMLASS	Concentration of active solids held in the aerator, mg/l
XMLBSS	Concentration of unmetabolized biodegradable solids held in aerator, mg/l
XMLDSS	Concentration of non-biodegradable solids in aerator caused by destruction of active solids by natural causes, mg/l
XMLISS	Concentration of inert inorganic solids in aerator, caused by inorganic solids in the influent, mg/l
XMLSS	Total concentration of solids in the aerator, mg/l
XMU	Dynamic viscosity of filtrate of sand drying beds, gm/cm-sec
XRSS	Ratio of solids in the overflow stream from the aerator to the total solids held, TSS(5)/MLSS
Y1	Alkalinity of sludge expressed as a percentage of the total solids
Y2	Ratio of volatile matter to ash for sludge.
YTONS	Number of tons of sludge incinerated

APPENDIX III

Program Listing in Fortran IV

```
01000↑PROGRAM SEWAGE
1020↑ DIMENSION NFORK1(15),UNIT(30),OUTPUT(20)
01030↑COMMON     Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
01040CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
01050CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
01060CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
01070↑COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
01080CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
01090↑COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
01100↑COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
01110CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
1111↑ COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
1112↑ COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
1115↑ INTEGER UNIT,OUTPUT,AEROBIC,DIGESTO
01120↑NCASE = 0
01130↑DO 300 I=1,20
01140↑Q(I)=0.0
01150↑SOC(I)=0.0
01160↑SNBC(I)=0.0
01170↑SON(I)=0.0
01180↑SOP(I)=0.0
01190↑SFM(I)=0.0
01200↑DOC(I)=0.0
01210↑DNBC(I)=0.0
01220↑DN(I)=0.0
01230↑DP(I)=0.0
01240↑DFM(I)=0.0
01250↑SBOD(I)=0.0
01260↑DBOD(I)=0.0
01270↑VSS(I)=0.0
01280↑TSS(I)=0.0
01290↑300 ALK(I)=0.0
01300↑DO 400 I=1,15
01310↑CCOST(I)=0.0
01320↑COSTO(I)=0.0
01330↑ACOST(I)=0.0
01340↑AOCOST(I)=0.0
01350↑400 CPERKG(I)=0.0
01360↑DO 500 I=1,12
01370↑500 ASS(I)=0.0
01380↑DO 600 I=1,10
01390↑NFORK(I)=0.0
01400↑600 DEMBOD(I)=0.0
01420↑102 FORMAT(4F8.4/6F10.5/5F10.4/5F10.4/5F10.4)
01430↑103 FORMAT(F8.2)
01440↑107  FORMAT(I3/(9F8.2))
01450↑301 FORMAT(1H1/7X,8HCASE NO.,I4,3X,
01460C36HACTIVATED SLUDGE PROCESS CALCULATION,2X,9HJUNE 1969/)
```

```
1470↑ 302 FORMAT(8X,7HSTATION,4X,3HMGD,4X,6HCARBON,7X,3HBOD,
1480C 6X,10HNBIOCARBON,3X,8HNITROGEN//)
1490↑ 323 FORMAT(//8X,7HSTATION,4X,3HMGD,4X,10HPHOSPHORUS,4X
1491C 5HFIXED,7X,3HVSS,7X,3HTSS/40X,6HMATTER/)
1510↑ 304 FORMAT(12X,3HDIS,8X,2(2X,E9.4),2(3X,E9.4)/)
1725↑ 156 FORMAT(/7X,I3,2X,3HSOL,3X,F6.3,1X,4E11.4)
1730↑108 FORMAT(8X,10H STRUCTURE,12X,8HCAP COST,5X,6HAMORT.,7X,
1740C9HOPERATION/30X,7HDOLLARS,6X,*$/YEAR*,7X,*$/YEAR*/)
1750↑109 FORMAT(//8X10H STRUCTURE,18X,*AMORTIZATION PLUS OPERATING COST*
1760C/37X*$/YEAR*10X,*CENTS/1000 GAL*/)
1775↑ 159 FORMAT(12X,3HDIS,10X,2E11.4)
1780↑155 FORMAT(8X,2A8,4X,3(E10.4,3X))
1790↑158 FORMAT(8X,2A8,12X,E10.4,6X,E10.4)
1800↑157 FORMAT(10A6/10A6/10A6)
1900↑ 610 FORMAT(7X,* MLASS = *E10.4,3X* MLBSS = *E10.4,3X,
1902C *MLNBSS = *E10.4/6X*  MLDSS = *E10.4,3X,*  MLSS = *E10.4,
1904C 3X,*  VAER = *E10.4/6X*   VNIT = *E10.4,3X,*RETURN = *
1906C E10.4,3X,* MLISS = *E10.4/6X*    CWR = *E10.4,3X,*CARREM = *,
1908C E10.4,3X,*BODREM = *E10.4/)
1910↑ 620 FORMAT(7X,* NITREM = *E10.4,2X,*PHOSREM = *E10.4,
1912C 2X,*  FRPS = *E10.4,/6X*   URPS = *E10.4,3X,*  URSS = *
1914C E10.4,3X,*  XRSS = *E10.4/6X*    EFF = *E10.4,3X,
1916C *   GPS = *E10.4,3X,*  GSTH = *E10.4/6X*   ATHM = *E10.4,
1918C 3X,*   TRR = *E10.4,3X,*    GE = *E10.4/)
1920↑ 630 FORMAT(6X*    GES = *E10.4,3X,*    AE = *E10.4,3X,
1922C *   ERR = *E10.4/6X*    WRE = *E10.4,3X,*  TDIG = *E10.4,
1924C 3X,*    TD = *E10.4/6X*   VDIG = *E10.4,3X,* FRDIG = *,
1926C E10.4,3X,* C1DIG = *E10.4/6X*  C2DIG = *E10.4,3X,*CH4CFD = *,
1928C E10.4,3X,*CO2CFD = *E10.4/)
1930↑ 640 FORMAT(9X,*VFL = *E10.4,3X,*   TVF = *E10.4,3X,*   AVF = *,
1931C E10.4/6X*   FOOD = *E10.4,3X,*  DEGC = *E10.4,3X,*    DO = *,
1933C E10.4/)
1940↑ 641 FORMAT(6X* VOLDIG = *,E10.4,3X,* ARATE = *,E10.4,3X,
1941C *AYEARS = *E10.4/6X* SOLIDS = *,E10.4,3X,* TSSLA = *,E10.4
1942C ,3X,*  DIGT = *E10.4//6X*  CPTON = *,E10.4)
02100↑110 FORMAT(8X,18HTOTAL CAPITAL COST,27X,E10.4
2107C/8X,*TOTAL AMORTIZATION AND OPERATING COST*8X,E10.4/8X,
2108C*AMORTIZATION COST PER 1000 GAL*15X,E10.4/8X,
2109C*OPERATING COST PER 1000 GAL*18X,E10.4/8X,*TOTAL COST PER*
2110C* 1000 GAL*22X,E10.4//)
02130↑306 FORMAT(6X,30HBOD5 DEMAND CANNOT BE ACHIEVED)
02190↑307 FORMAT(6X,27HDEMAND MLASS CANNOT BE HELD)
2200↑ 313 FORMAT(7X,I3,2X,3HSOL,2X,F6.3,2(1X,E10.4),2(2X,E10.4))
02210↑308 FORMAT(7X,5HECF  ,7F5.2/8X,8F5.2)
02250↑305 FORMAT (15F4.2)
02260↑700 FORMAT(7X,7HCENG = E10.4,2X,7HCTRP = E10.4,2X,7HCTGO = E10.4,
2270C/7X,8HCLAND = E10.4,2X,6HCCR = E10.4)
```

```
02280↑701 FORMAT(7X,6HCCI = E10.4,6X,5HAF = E10.4)
02290↑325 FORMAT(1H1)
2310↑321 FORMAT(15X,8A5/15X,8A5)
2315↑ 324 FORMAT(///)
02320↑322  FORMAT(//10X17H INPUT PARAMETERS,///15X17H SOLID BOD (MG/L),2
02330C0X,
02340CF8.2//16X20HDISSOLVED BOD (MG/L)16X,F8.2//15X30H TOTAL SUSPENDED S
02350COLIDS (MG/L)7XF8.2//15X33H VOLATILE SUSPENDED SOLIDS (MG/L)4XF8.2/
02360C/15X20H SEWAGE VOLUME (MGD)17XF8.2//15X30H MIXED LIQUOR SUSPENDED
02370CSOLIDS/15X23H HELD IN AERATOR (MG/L)14XF8.2//15X31H MAXIMUM BOD IN
2380C EFFLUENT (MG/L)6XF8.2/1H0)
2390↑ READ,(UNIT(II),II=1,30)
2392↑ READ,AEROBIC,DIGESTO
2395↑ READ,(OUTPUT(I),I=1,7)
2396↑ 2396 FORMAT(2A8)
02400↑READ, (ECF(I),I=1,15)
02410↑READ, (NFORK(I),I=1,10)
2411↑ IF(NFORK(2) .EQ. 1) NFORK1(2) = 1
2412↑IF(NFORK(3) .EQ.0) NFORK1(8) = 1
2413↑IF(NFORK(3) .EQ.0) NFORK1(10) = 1
2414↑IF(NFORK(1) .EQ.1) NFORK1(11) = 1
2415↑IF(NFORK(1) .EQ.0) NFORK1(13) = 1
2416↑ IF(NFORK(1) .EQ. 1) NFORK1(10) = 1
2417↑ IF(NFORK(3) .EQ. 0) UNIT(17) = AEROBIC
2418↑ IF(NFORK(3) .EQ.0) UNIT(18) = DIGESTO
02420↑READ, SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
02430CDNBC(20),DN(20),DP(20),DFM(20),DEGC,ALK(20)
02440↑VSS(20)=SOC(20)*2.13
02450↑TSS(20)=VSS(20)+SFM(20)
02460↑SBOD(20)=(SOC(20)-SNBC(20))*1.87
02470↑DBOD(20)=(DOC(20)-DNBC(20))*1.87
02480↑READ , URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,GSS,
02490CTRR,TSS(12),GTH,GSTH,ERR,TSS(15),WRE,GE,GES,TDIG,TD,
02500CTSS(16),VFL,TVF,SBL
2505↑ READ,AYEARS,ARATE
2510↑ AF = (ARATE*(1. + ARATE)**AYEARS)/((ARATE + 1.)**AYEARS -1.)
02530↑CCL2=0.08
02540↑DCL2=8.0
02550↑READ , NAS,(ASS(I),I=1,NAS)
02560↑READ, NBOD,(DEMBOD(J),J=1,NBOD)
02570↑READ,NQ,(Q20(J),J=1,NQ)
02580↑READ, NFR,(FRPSIN(J),J=1,NFR)
2581↑ READ,XMU,HC,SIGMA,HO,SAND,DRYDAY,XLANDC
2582↑ READ,SCI,STO,STF,RC
2587↑LOOPS = 6
2598 CONTINUE
```

```
DO 25 LF=1,NFR
FRPS=FRPSIN(LF)
DO 25 LQ=1,NQ
Q(20)=Q20(LQ)
80 DO 25 I=1,NAS
XMLSS=ASS(I)
DO 25 K=1,NBOD
BOD5=DEMBOD(K)
67   CONTINUE
CALL BIOLOG
IF(K35 .EQ. 5) WRITE(61,307)
IF(K35 .EQ. 10) WRITE(61,306)
 IF(K35 .EQ. 5 .OR. K35 .EQ. 10) GO TO 25
IF(NFORK(3) .EQ. 0) GO TO 81
CALL THICK
CONTINUE
 GO TO 82
81 CALL AEROBE
GO TO 83
82 CALL DIGEST
83 CONTINUE
 IF(NFORK(1)) 77,77,78
 78 CALL DRY
 GO TO 46
 77 CONTINUE
IF(NFORK(1).EQ. 1) GO TO 46
CALL VACUUM
CONTINUE
 46 CONTINUE
Q(9)=Q(11)+Q(14)+Q(16)
TEMP1=Q(11)/Q(9)
TEMP2=Q(14)/Q(9)
TEMP3=Q(16)/Q(9)
SOC(9)=TEMP1*SOC(11)+TEMP2*SOC(14)+TEMP3*SOC(16)
SNBC(9)=TEMP1*SNBC(11)+TEMP2*SNBC(14)+TEMP3*SNBC(16)
SON(9)=TEMP1*SON(11)+TEMP2*SON(14)+TEMP3*SON(16)
SOP(9)=TEMP1*SOP(11)+TEMP2*SOP(14)+TEMP3*SOP(16)
SFM(9)=TEMP1*SFM(11)+TEMP2*SFM(14)+TEMP3*SFM(16)
DOC(9)=TEMP1*DOC(11)+TEMP2*DOC(14)+TEMP3*DOC(16)
DNBC(9)=TEMP1*DNBC(11)+TEMP2*DNBC(14)+TEMP3*DNBC(16)
DN(9)=TEMP1*DN(11)+TEMP2*DN(14)+TEMP3*DN(16)
DP(9)=TEMP1*DP(11)+TEMP2*DP(14)+TEMP3*DP(16)
DFM(9)=TEMP1*DFM(11)+TEMP2*DFM(14)+TEMP3*DFM(16)
SBOD(9)=(SOC(9)-SNBC(9))*1.87
DBOD(9)=(DOC(9)-DNBC(9))*1.87
SBOD(9)= (SOC(9)-SNBC(9))*1.87
```

```
VSS(9)=SOC(9)*2.0
TSS(9)=VSS(9)+SFM(9)
Q(4)=Q(2)+Q(6)
*      CALCULATE CAPITAL AND OPERATING COST
 CCOST(1)=14700.*Q(20)**0.625
COSTO(1)=500.0*Q(20)+2150.0*Q(20)**0.37
GPS=-2780.0*ALOG(FRPS)-551.7
 IF(NFORK(2) .EQ. 1) GO TO 1001
APS=Q(1)*1000.0/GPS
APS=APS*ECF(2)
CCOST(2)=13400.0*APS+5200.0*APS**0.10
COSTO(2)=1000.0*APS+2500.0*APS**0.5
1001 VAER=VAER*ECF(3)
CCOST(3)=175000.0*VAER+36500.0*VAER**0.182
 COSTO(3) =10000.*(VAER+VOLDIG) + 14500.*(VAER+VOLDIG)**0.5
BSIZE=BSIZE*ECF(4)
CCOST(4)=10570.0+5.857*BSIZE
CAIRP=AIRCFD*365.0*CKWH/1830.0
COSTO(4)=CAIRP
 Q(4) = Q(2) + Q(6)
AFS=Q(4)*1000.0/GSS
AFS=AFS*ECF(5)
CCOST(5)=12600.0*AFS+5350.0/AFS**0.126
Q(6)=Q(6)*ECF(6)
CCOST(6)=3650.0+2250.0*Q(6)
CCOST(7)=40000.0*Q(20)**0.70
ATHM=ATHM*ECF(8)
CCOST(8)=(18.8+9.1/EXP(ATHM/13300.0))*ATHM
VDIG=VDIG*ECF(9)
CCOST(9)=5000.0+1080.0*VDIG+10700.0*VDIG**0.128
 IF(NFORK(3) .EQ. 0)CCOST(9)=175000.*VOLDIG+36500.*VOLDIG**.182
AE=AE*ECF(10)
 CCOST(10) = (18.8+52.0/EXP(AE/6000.))*AE
AVF=AVF*ECF(11)
CCOST(11)=12800.0+372.0*AVF
 TONS = TSS(18)*365./2000.
COSTO(11)=TSS(15)*FECL3*Q(15)*30.40*CFECL3
COSTO(11)=COSTO(11)+1500.0*Q(20)+6450.0*Q(20)**0.37
YTONS=0.1825*TSS(18)
COSTO(12)=16.1*YTONS-0.00009*YTONS**2.
CCOST(12)=1570.0*TSS(18)**0.60
CCOST(14)=9000.0*(Q(5)-Q(17))**0.469
COSTO(14)=DCL2*(Q(5)-Q(17))*8.33*CCL2*365.0
CCOST(15)=4400.0*Q(20)**0.875
IF(NFORK(1))48,48,47
 47 ASB = AREA
CCOST(10)=0.0
```

```
 IF(NFORK(3) .EQ. 0) VDIG = Q(13)/2.5E-4
48 COSTO(9)=48.0*VDIG+540.0*VDIG**0.44
 IF(NFORK(3) .EQ. 0) COSTO(9) = 0.0
  49  CTRP=0.10
CTGO=0.15
CLAND=0.02
 DO 30 J=1,15
30 CCOST(J)=CCOST(J)*CCI
DO 31 J=1,15
31 ACOST(J)=CCOST(J)*AF
TOTCC=0.0
DO 32 J=1,15
32 TOTCC =TOTCC+CCOST(J)
CENG=0.08*(1000000.0/TOTCC)**0.146
CCR=1.0+CENG+CTRP+CTGO+CLAND
TOTCC=TOTCC*CCR
DO 90 J=1,15
ACOST(J)=ACOST(J)*CCR
90 CCOST(J)=CCOST(J)*CCR
TACOST=0.0
TCOSTO=0.0
TOTAOC=0.0
DO 33 J=1,15
TACOST=TACOST+ACOST(J)
AOCOST(J)=ACOST(J)+COSTO(J)
CPERKG(J)=AOCOST(J)/Q(20)/3650.0
TOTAOC=TOTAOC+AOCOST(J)
33 TCOSTO=TCOSTO+COSTO(J)
 CPTON = AOCOST(11)/TONS
CAPKG=TACOST/Q(20)/3650.0
COPKG=TCOSTO/Q(20)/3650.0
TCOST=TOTAOC/Q(20)/3650.0
TOC2=SOC(2)+DOC(2)
TOC5=SOC(5)+DOC(5)
TOC7=SOC(7)+DOC(7)
TBOD2=SBOD(2)+DBOD(2)
TBOD5=SBOD(5)+DBOD(5)
CWR=Q(7)*TOC7/Q(2)/TOC2
CREM=1.0-TOC5*Q(5)/Q(2)/TOC2
BODREM=1.0-TBOD5*Q(5)/Q(2)/TBOD2
EFF=CREM-CWR
REMN=1.0-(SON(5)+DN(5))*Q(5)/Q(2)/(SON(2)+DN(2))
REMP=1.0-(SOP(5)+DP(5))*Q(5)/Q(2)/(SOP(2)+DP(2))
LOOPS=LOOPS-1
IF(LOOPS)67,68,67
68    NCASE = NCASE + 1
 IF(OUTPUT(1) .EQ.1) GO TO 604
```

```
03910 PRINT 322, SBOD(20),DBOD(20),TSS(20),VSS(20),Q(20),XMLSS,BODS
3930↑ 604 IF( OUTPUT(2) .EQ.1) GO TO 601
3935 PRINT 301,NCASE
03940 PRINT 302
03950↑NSTA=1
3960↑ DO 91 K7 = 1,20
3970↑IF(Q(K7) .LE. 0.0001) GO TO 91
3975↑ IF(SOC(K7) .LE. 0.001) GO TO 91
3980↑PRINT 313,K7,Q(K7),SOC(K7),SBOD(K7),SNBC(K7),SON(K7)
3990↑PRINT 304,DOC(K7),DBOD(K7),DNBC(K7),DN(K7)
4000↑ 91 CONTINUE
4006↑ PRINT 323
4010↑DO 92 K8=1,20
4020↑IF(Q(K8) .LE. 0.0001) GO TO 92
4025↑ IF( SOC(K8) .LE. 0.001) GO TO 92
4030↑PRINT 156,K8,Q(K8),SOP(K8),SFM(K8),VSS(K8),TSS(K8)
4040↑PRINT 159,DP(K8),DFM(K8)
4050↑ 92 CONTINUE
4060↑ 601 IF(OUTPUT(3) .EQ. 1) GO TO 4800
4600↑ PRINT 301,NCASE
4690↑ PRINT 108
4700↑DO 93 II=1,15
4710↑IF(NFORK1(II) .EQ.1) GO TO 93
4715↑ I7 = 2*II - 1
4720↑PRINT155,UNIT(I7),UNIT(I7+1),CCOST(II),ACOST(II),COSTO(II)
4730↑93 CONTINUE
4735↑ PRINT 109
4740↑DO 94 II=1,15
4750↑IF(NFORK1(II) .EQ. 1) GO TO 94
4755↑ I7= 2*II-1
4760↑PRINT158,UNIT(I7),UNIT(I7+1),AOCOST(II),CPERKG(II)
4770↑ 94 CONTINUE
4780↑ PRINT 324
4800 CONTINUE
04850 PRINT 110, TOTCC,TOTAOC,CAPKG,COPKG,TCOST
4855↑ 602 IF(OUTPUT(4) .EQ. 1) GO TO 603
4860↑ PRINT 610,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLSS,VAER,VNIT,
4870C RETURN,XMLISS,CWR,CREM,BODREM
4880↑ PRINT 620, REMN,REMP,FRPS,URPS,URSS,XRSS,EFF,
4890C GPS,GSTH,ATHM,TRR,GE
4900↑ PRINT 630,GES,AE,ERR,WRE,TDIG,TD,VDIG,FRDIG,C1DIG,
4910C C2DIG,CH4CFD,CO2CFD
4920↑ PRINT 640, VFL,TVF,AVF,FOOD,DEGC,DO
4930↑ PRINT 641, VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT,CPTON
04950 PRINT 308, (ECF(J),J=1,15)
04960 PRINT 700, CENG,CTRP,CTGO,CLAND,CCR
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```
04970 PRINT 701,  CCI,AF
4974↑ 603 CONTINUE
4980↑ PRINT 325
04990↑LOOPS=6
05000↑25   CONTINUE
05010↑STOP
05020↑END
```

```
SUBROUTINE BIOLOG
COMMON     Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
 COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
 COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
67 IF(LOOPS-6)66,65,66
*      MIX STREAMS NINE AND TWENTY
65 Q(9)=0.0
 FMIN = 0.0
66 TEMP1=Q(20)/(Q(20)+Q(9))
TEMP2=Q(9)/(Q(20)+Q(9))
SOC(1)=TEMP1*SOC(20)+TEMP2*SOC(9)
SON(1)=TEMP1*SON(20)+TEMP2*SON(9)
SOP(1)=TEMP1*SOP(20)+TEMP2*SOP(9)
SNBC(1)=TEMP1*SNBC(20)+TEMP2*SNBC(9)
SFM(1)=TEMP1*SFM(20)+TEMP2*SFM(9)
DOC(1)=TEMP1*DOC(20)+TEMP2*DOC(9)
DNBC(1)=TEMP1*DNBC(20)+TEMP2*DNBC(9)
DN(1)=TEMP1*DN(20)+TEMP2*DN(9)
DP(1)=TEMP1*DP(20)                +TEMP2*DP(9)
DFM(1)=TEMP1*DFM(20)+TEMP2*DFM(9)
ALK(1)=TEMP1*ALK(20)+TEMP2*ALK(9)
Q(1)=Q(20)+Q(9)
VSS(1)=SOC(1)*2.13
TSS(1)=VSS(1)+SFM(1)
SBOD(1)=(SOC(1)-SNBC(1))*1.87
 DBOD(1) = (DOC(1) -DNBC(1))*1.87
IF(NFORK(2)) 100,100,110
100  CALL PRIMARY
GO TO 120
110  SOC(2)=TEMP1*SOC(20)+TEMP2*SOC(9)
SON(2)=TEMP1*SON(20)+TEMP2*SON(9)
SOP(2)=TEMP1*SOP(20)+TEMP2*SOP(9)
SNBC(2)=TEMP1*SNBC(20)+TEMP2*SNBC(9)
SFM(2)=TEMP1*SFM(20)+TEMP2*SFM(9)
DOC(2)=TEMP1*DOC(20)+TEMP2*DOC(9)
DNBC(2)=TEMP1*DNBC(20)+TEMP2*DNBC(9)
DN(2)=TEMP1*DN(20)+TEMP2*DN(9)
DP(2)=TEMP1*DP(20)                +TEMP2*DP(9)
```

```
DFM(2)=TEMP1*DFM(20)+TEMP2*DFM(9)
ALK(2)=TEMP1*ALK(20)+TEMP2*ALK(9)
Q(2)=Q(20)+Q(9)
VSS(2)=SOC(1)*2.13
TSS(2)=VSS(1)+SFM(1)
SBOD(2)=(SOC(1)-SNBC(1))*1.87
DBOD(2)=(DOC(1)-DNBC(1))*1.87
 120 CONTINUE
 BOD2 = SBOD(2) + DBOD(2)
DBOD2=DBOD(2)
*      AERATOR PERFORMANCE
1000 CEDR=0.18*1.047**(DEGC-28.0)
CAER=CAER20*1.047**(DEGC-20.0)
SVI=100.0
SMAX=1200000.0/SVI
 IF(NFORK(2).EQ.1) SMAX = 3000000./SVI
IF (URSS*XMLSS-SMAX) 86,86,85
85 URSS=SMAX/XMLSS
86 SA=XMLSS/1000.0
TA=(BOD2-BOD5)/(BOD5*CAER*SA*24.)
 VAER = Q(2)*TA
XRSS=556.1*GSS**0.4942/XMLSS**1.8165/(TA*24.)**0.4386
ASMAX=BOD5/XRSS/0.685
ASMIN=0.0
IF(ASMAX-XMLSS)50,50,70
70   ASMAX = XMLSS
50 XMLASS=(ASMAX+ASMIN)/2.0
42 FOOD=SBOD(2)+DBOD2
FMAX=FOOD
N=1
GO TO 8
7 ERROR=FMAX-FMIN
TOL=0.10
IF(ERROR-TOL)21,21,19
19 FOOD=(FMIN+FMAX)/2.0
4 IF(FOOD-DBOD(2))5,5,6
5 DBOD(4)=DBOD(2)-FOOD
SBOD(4)=SBOD(2)
GO TO 8
6 SBOD(4)=(SBOD(2)+DBOD(2)-FOOD)*0.70
DBOD(4)=0.233*SBOD(4)
8 TEMP1=(0.65*FOOD/XMLASS)-XRSS
Q(7)=(Q(2)*TEMP1-CEDR*VAER)/(URSS-XRSS)
20 Q(5)=Q(2)-Q(7)
N=N+1
IF(N-2)23,23,22
22 TEMP2=XRSS*Q(5)+URSS*Q(7)
XMLBSS=Q(2)*SBOD(4)/TEMP2/0.80
```

```
SBOD(5)=(XMLASS*0.685+XMLBSS*0.80)*XRSS
DBOD(5)=DBOD(4)
TBOD5=SBOD(5)+DBOD(5)
GO TO 24
23 TBOD5=XMLASS*XRSS*0.685
24 IF(TBOD5-BOD5)10,10,15
10 IF(N-3)11,12,13
11 FMIN=(CEDR*VAER/Q(2)+XRSS)*XMLASS/0.65
FOOD=FMIN
GO TO 4
12   K35 = 5
RETURN
13 FMAX=FOOD
GO TO 7
15 IF(N-3)16,18,17
16   K35 = 10
RETURN
17 FMIN=FOOD
18 GO TO 7
21 CONTINUE
XLNBSS=SNBC(2)*Q(2)*2.13/TEMP2
XMLISS=Q(2)*SFM(2)/TEMP2
XMLDSS=(0.12*Q(2)*FOOD/TEMP2)-0.185*XMLASS
TEMP1=XMLASS+XMLBSS+XLNBSS+XMLISS+XMLDSS
TEMP2=TEMP1-XMLSS
TOL=5.0
IF(ABS(TEMP2)-TOL)41,41,51
51 IF(TEMP2)52,52,53
52 ASMIN=XMLASS
GO TO 50
53 ASMAX=XMLASS
GO TO 50
41 CONTINUE
SOC(5)=(XMLDSS+XMLASS)*XRSS/2.46+(XMLBSS+XLNBSS)*XRSS/2.33
SOC(7)=SOC(5)*URSS/XRSS
SNBC(5)=XLNBSS*XRSS/2.33+(XMLDSS+0.185*XMLASS)*XRSS/2.46
SNBC(7)=SNBC(5)*URSS/XRSS
TEMP2=XRSS*XMLASS/2.46
SON(5)=0.234*TEMP2+(SOC(5)-TEMP2)/10.0
TEMP2=TEMP2*URSS/XRSS
SON(7)=0.234*TEMP2+(SON(7)-TEMP2)/10.0
SOP(5)=SOC(5)*0.01
SOP(7)=SOC(7)*0.01
SFM(5)=XMLISS*XRSS
SFM(7)=XMLISS*URSS
DOC(5)=DNBC(2)+DBOD(4)/1.87
```

```
DNBC(5)=DNBC(2)
DOC(7)=DOC(5)
DNBC(7)=DNBC(5)
DN(5)=(Q(2)*(SON(2)+DN(2))-(SON(5)*Q(5)+SON(7)*Q(7)))/(Q(5)+Q(7))
DN(7)=DN(5)
DP(5)=(Q(2)*(SOP(2)+DP(2))-(SOP(5)*Q(5)+SOP(7)*Q(7)))/(Q(5)+Q(7))
DP(7)=DP(5)
DFM(5)=DFM(2)
DFM(7)=DFM(2)
SBOD(7)=(SOC(7)-SNBC(7))*1.87
DBOD(7)=(DOC(7)-DNBC(7))*1.87
VSS(5)=SOC(5)*1.90
TSS(5)=VSS(5)+SFM(5)
VSS(7)=SOC(7)*1.90
TSS(7)=VSS(7)+SFM(7)
*     CONDITIONS FOR NITRIFICATION
Q(6)=(Q(2)*(1.0-0.65*FOOD/XMLASS)+CEDR*VAER)/(URSS-1.0)
RETURN=Q(6)/Q(2)
X4X3=(1.0+RETURN)/RETURN/URSS
DN(3)=DN(2)/(1.0+RETURN)
X3Y=DN(3)*0.99/(X4X3-1.0)
CNIT=0.18*EXP(0.116*(DEGC-15.0))
TAN=(1.0+RETURN)*(ALOG(X4X3)+4.605/(DN(3)+X3Y))/CNIT
VNIT=Q(2)*TAN
*      AIR REQUIREMENTS
DOSAT=14.16-0.3943*DEGC+0.007714*DEGC**2-0.0000646*DEGC**3
DOSAT=DOSAT*1.221
AEFF=AEFF20*(DOSAT-DO)*1.02**(DEGC-20.0)/DOSAT
WFOOD=Q(2)*FOOD*8.33
WAS=XMLASS*VAER*8.33
AIRCFD=(0.577*WFOOD+1.16*CEDR*WAS)/AEFF/0.232/0.075
WDN=(Q(5)*DN(5)+Q(7)*DN(7))*8.33
IF(VNIT-VAER)26,26,27
26 AIRCFD=AIRCFD+4.6*WDN/AEFF/0.232/0.075
27 BSIZE=AIRCFD/1440.0
CFPGAL =AIRCFD/1000000.0/Q(1)
Q(10)=Q(7)+Q(8)
SOC(10)=(SOC(7)*Q(7)+SOC(8)*Q(8))/Q(10)
SNBC(10) =(SNBC(7)*Q(7)+SNBC(8)*Q(8))/Q(10)
SON(10)=(SON(7)*Q(7)+SON(8)*Q(8))/Q(10)
SOP(10)=(SOP(7)*Q(7)+SOP(8)*Q(8))/Q(10)
SFM(10)=(SFM(7)*Q(7)+SFM(8)*Q(8))/Q(10)
DOC(10)=(DOC(7)*Q(7)+DOC(8)*Q(8))/Q(10)
DNBC(10)=(DNBC(7)*Q(7)+DNBC(8)*Q(8))/Q(10)
DN(10)=(DN(7)*Q(7)+DN(8)*Q(8))/Q(10)
DP(10)=(DP(7)*Q(7)+DP(8)*Q(8))/Q(10)
DFM(10)=(DFM(7)*Q(7)+DFM(8)*Q(8))/Q(10)
```

```
06860↑VSS(10)=SOC(10)*2.0
06870↑TSS(10)=VSS(10)+SFM(10)
06880↑RETURN
06890↑END


06900↑SUBROUTINE THICK
06910↑COMMON     Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
06920CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
06930CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
06940CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
06950↑COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
06960CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
06970↑COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
06980↑COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
06990CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
6995↑ COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
6996↑ COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
07000*      CALCULATE STREAM INTO THICKNER
07010*      CALCULATE STREAMS FROM THICKNER
07020↑Q(12)=TRR*Q(10)*TSS(10)/TSS(12)
07030↑Q(11)=Q(10)-Q(12)
07040↑TSS(11)=(1.0-TRR)*Q(10)*TSS(10)/Q(11)
07050↑TEMP4=TSS(11)/TSS(10)
07060↑SOC(11)=TEMP4*SOC(10)
07070↑VSS(11)=SOC(11)*2.0
07080↑SNBC(11)=TEMP4*SNBC(10)
07090↑SON(11)=TEMP4*SON(10)
07100↑SOP(11)=TEMP4*SOP(10)
07110↑SFM(11)=TEMP4*SFM(10)
07120↑DOC(11)=DOC(10)
07130↑DNBC(11)=DNBC(10)
07140↑DN(11)=DN(10)
07150↑DP(11)=DP(10)
07160↑DFM(11)=DFM(10)
07170↑TEMP3=TSS(12)/TSS(10)
07180↑ATH2=Q(10)*TSS(10)*8.33/GSTH
07190↑SOC(12)=TEMP3*SOC(10)
07200↑SNBC(12)=TEMP3*SNBC(10)
07210↑SON(12)=TEMP3*SON(10)
07220↑SOP(12)=TEMP3*SOP(10)
07230↑SFM(12)=TEMP3*SFM(10)
07240↑DOC(12)=DOC(10)
07250↑DNBC(12)=DNBC(10)
```

```
07260↑DN(12)=DN(10)
07270↑DP(12)=DP(10)
07280↑DFM(12)=DFM(10)
07290↑VSS(12)=SOC(12)*2.0
07300↑ATH1=Q(10)*1000000.0/GTH
07310↑ATHM=AMAX1(ATH1,ATH2)
07320↑RETURN
07330↑END
```

```
01000↑SUBROUTINE DIGEST
01010↑COMMON    Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
01020CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
01030CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
01040CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
01050↑COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
01060CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
01070↑COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
01080↑COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
01090CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
1095 COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
1096↑ COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
01100*     CALCULATE DIGESTER PERFORMANCE
01110↑C1DIG=0.28/EXP(0.036*(35.0-TDIG))
01120↑C2DIG=700.00*EXP(0.10*(35.0-TDIG))
01130↑DIGC12=SOC(12)-SNBC(12)+DOC(12)-DNBC(12)
01140↑TDM=2.5/C1DIG
01150↑TDOPT=(1.0-(C2DIG/(C2DIG+DIGC12)**0.5))/C1DIG
01160↑TDMIN=AMAX1(TDM,TDOPT)
01170↑IF(TD-TDMIN)43,43,44
01180↑43 TD=TDMIN
01190↑44 DIGC13=C2DIG/(C1DIG*TD-1.0)
01200↑TEMP4=(DIGC12-DIGC13)/(SOC(12)+DOC(12))
01210↑CDF=200.00*EXP(0.12*(35.0-TDIG))
01220↑DF=CDF/(C1DIG*TD-1.0)
01230↑SOC(13)=SNBC(12)+DIGC13-DF
01240↑DOC(13)=DNBC(12)+DF
01250↑FRDIG=(SOC(12)-SOC(13))/SOC(12)
01260↑SNBC(13)=SNBC(12)
01270↑SON(13 )=(1.0-TEMP4)*SON(12)
01280↑SOP(13)=(1.0-TEMP4)*SOP(12)
01290↑SFM(13)=SFM(12)
01300↑DNBC(13)=DNBC(12)
01310↑DN(13)=DN(12)+0.65*SON(12)*TEMP4
01320↑DP(13)=DP(12)+SOP(12)*TEMP4
01330↑DFM(13)=DFM(12)
01340↑VDIG=Q(12)*TD*1000.0/7.48
```

```
CH4CFD=163.85*(DIGC12-DIGC13)*Q(12)
CO2CFD=249.9*(DIGC12-DIGC13)*Q(12)-CH4CFD
VSS(13)=SOC(13)*2.0
TSS(13)=VSS(13)+SFM(13)
Q(13)=Q(12)
*      CALCULATE STREAMS FROM SLUDGE WASH
IF(NFORK(1))45,45,46
46    RETURN
45 Q(17)=WRE*Q(13)
TEMP1=Q(13)*1000000.0/GE
TEMP2=Q(13)*TSS(13)*8.33/GES
AE=AMAX1(TEMP1,TEMP2)
Q(15)=ERR*Q(13)*TSS(13)/TSS(15)
Q(14)=Q(17)+Q(13)-Q(15)
TSS(14)=Q(13)*((1.0-ERR)*TSS(13)+WRE*TSS(5))/Q(14)
VSS(17)=VSS(5)
TSS(17)=TSS(5)
TEMP1=TSS(15)/TSS(13)
SOC(15)=TEMP1*SOC(13)
SNBC(15)=TEMP1*SNBC(13)
SON(15)=TEMP1*SON(13)
SOP(15)=TEMP1*SOP(13)
SFM(15)=TEMP1*SFM(13)
VSS(15)=SOC(15)*2.0
TEMP2=TSS(14)/TSS(13)
SOC(14)=TEMP2*SOC(13)
VSS(14)=SOC(14)*2.0
SNBC(14)=TEMP2*SNBC(13)
SON(14)=TEMP2*SON(13)
SOP(14)=TEMP2*SOP(13)
SFM(14)=TEMP2*SFM(13)
TEMP1= Q(13)/(Q(13)+Q(17))
TEMP2=Q(17)/(Q(13)+Q(17))
DOC(14)=TEMP1*DOC(13)+TEMP2*DOC(5)
DNBC(14)=TEMP1*DNBC(13)+TEMP2*DNBC(5)
DN(14)=TEMP1*DN(13)+TEMP2*DN(5)
DP(14)=TEMP1*DP(13)+TEMP2*DP(5)
DFM(14)=TEMP1*DFM(13)+TEMP2*DFM(5)
DOC(15)=DOC(14)
DNBC(15)=DNBC(14)
DN(15)=DN(14)
DP(15)=DP(14)
DFM(15)=DFM(14)
IF(VNIT-VAER)60,60,61
60 ALK(5)=ALK(1)+3.57*(DN(5)-DN(2))
GO TO 62
61 ALK(5)=ALK(1)
62 ALK(12)=ALK(5)
ALK(13)=ALK(12)+(DN(13)-DN(12))*3.57
ALK(14)=TEMP1*ALK(13)+TEMP2*ALK(5)
ALK(15)=ALK(14)
RETURN
END
```

```
SUBROUTINE VACUUM
COMMON     Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
 COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS
Y1=100.00*ALK(15)/TSS(15)
Y2=VSS(15)/(TSS(15)-VSS(15))
FECL3=1.08*Y1+2.0*Y2
CFECL3=0.08
*     CALCULATE STREAMS FROM VACUUM FILTER
WP=88.0/(TSS(15)/10000.0)**0.123
TEMP1=1000000.0*(100.0-WP)/WP
Q(16)=Q(15)*(TEMP1-TSS(15))/(TEMP1-TSS(16))
TSS(18)=8.33*(Q(15)*TSS(15)-Q(16)*TSS(16))
TEMP2=TSS(18)/TSS(15)
SOC(18)=TEMP2*SOC(15)
VSS(18)=SOC(18)*2.0
SNBC(18)=TEMP2*SNBC(15)
SON(18)=TEMP2*SON(15)
SOP(18)=TEMP2*SOP(15)
VSS(16)=SOC(16)*2.0
DOC(16)=DOC(15)
DNBC(16)=DNBC(15)
DN(16)=DN(15)
DP(16)=DP(15)
TEMP3 =8.33*(Q(15)-Q(16))
DFM(18)=DFM(15)*TEMP3
DP(18)=DP(15)*TEMP3
DN(18)=DN(15)*TEMP3
DNBC(18)=DNBC(15)*TEMP3
DOC(18)=DOC(15)*TEMP3
TEMP2=TSS(16)/TSS(15)
SOC(16)=SOC(15)*TEMP2
SNBC(16)=SNBC(15)*TEMP2
SON(16)=SON(15)*TEMP2
SOP(16)=SOP(15)*TEMP2
SFM(16)=SFM(15)*TEMP2
DFM(16)=DFM(15)
AVF=Q(16)*1000000./VFL/TVF/24.
RETURN
END
```

```
  3000↑SUBROUTINE PRIMARY
3010*     PRIMARY SETTLER PERFORMANCE
3020↑COMMON    Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
3030CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
3040CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
3050CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
3060↑COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
3070CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
3080↑COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
3090↑COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
3100CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
3110↑ COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS
3120↑Q(8)=FRPS*Q(1)/URPS
3130↑Q(2)=Q(1)-Q(8)
3140↑TEMP1=(1.-FRPS)*Q(1)/Q(2)
3150↑SOC(2)=TEMP1*SOC(1)
3160↑SNBC(2)=TEMP1*SNBC(1)
3170↑SON(2)=TEMP1*SON(1)
3180↑SOP(2)=TEMP1*SOP(1)
3190↑SFM(2)=TEMP1*SFM(1)
3200↑DOC(2)=DOC(1)
3210↑DNBC(2)=DNBC(1)
3220↑DN(2)=DN(1)
3230↑DP(2)=DP(1)
3240↑DFM(2)=DFM(1)
3250↑VSS(2)=SOC(2)*2.13
3260↑TSS(2)=VSS(2)+SFM(2)
3270↑SBOD(2)=(SOC(2)-SNBC(2))*1.87
3280↑DBOD(2)=(DOC(2)-DNBC(2))*1.87
3290↑TEMP1=FRPS*Q(1)/Q(8)
3300↑SOC(8)=TEMP1*SOC(1)
3310↑SNBC(8)=TEMP1*SNBC(1)
3320↑SON(8)=TEMP1*SON(1)
3330↑SOP(8)=TEMP1*SOP(1)
3340↑SFM(8)=TEMP1*SFM(1)
3350↑DOC(8)=DOC(2)
3360↑DNBC(8)=DNBC(2)
3370↑DN(8)=DN(2)
3380↑DP(8)=DP(2)
3390↑DFM(8)=DFM(2)
3400↑VSS(8)=SOC(8)*2.13
3410↑TSS(8)=VSS(8)+SFM(8)
3420↑SBOD(8)=(SOC(8)-SNBC(8))*1.87
3430↑DBOD(8)=(DOC(8)-DNBC(8))*1.87
3440↑RETURN
3450↑END
```

```
SUBROUTINE AEROBE
COMMON    Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS/VNIT
COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
 COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
 COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
 VOLAT = 60.
 DIGT = 7.0
 TSSLA = 0.01
 Q(12) = Q(10)
 SOLIDS = 6.0
 SOC(12) = SOC(10)
 SNBC(12) = SNBC(10)
 SON(12) = SON(10)
 SOP(12) = SOP(10)
 SFM(12) = SFM(10)
 TSS(12) = TSS(10)
 DOC(12) = DOC(10)
 VSS(12) = VSS(10)
 DNBC(12) = DNBC(10)
 DN(12) = DN(10)
 DP(12) = DP(10)
 DFM(12) = DFM(10)
 SBOD(12) = SBOD(10)
 DBOD(12) = DBOD(10)
* EFFECT OF THICKENING
 Q(13) = Q(12)*TSS(12)/(SOLIDS*10000.)
* EFFECT OF SOLIDS DESTRUCTION
 10 DIGT = DIGT + 1.
 VOLRED = 2.84 + 35.07*ALOG(DIGT)/2.3025
 VSSA = VSS(12)*(1. - VOLRED/100.)*(Q(12)/Q(13))
 TSSA = (TSS(12)-(VSS(12)*VOLRED/100.))*Q(12)/Q(13)
 IF(VOLRED .LE. VOLAT) GO TO 10
 Q(13) = Q(13)*TSSA/(SOLIDS*10000.)
 XRT = VSSA/TSSA
 TSSA = SOLIDS*10000.
 VSSA = XRT*TSSA
 Q(11) = Q(12) - Q(13)
 TSS(11) = TSSLA*SOLIDS*10000.
 TSS(13) = TSSA
 VSS(13) = VSSA*TSS(13)/(SOLIDS*10000.)
```

```
VSS(11) = TSSLA*VSSA
SOLIDNA = SON(12)*(1. -VOLRED/100.)*Q(12)/Q(13)
DNA = DN(12) + SON(12)*(VOLRED/100.)
DN(11) = DNA
DN(13) = DNA
SON(11) = SOLIDNA*TSSLA
SON(13) =Q(12)*SON(12)*(1.-TSSLA +VOLRED*TSSLA/100. -VOLRED/100.
)/Q(13)
SOLIDPA = SOP(12)*(1.-VOLRED/100.)*Q(12)/Q(13)
DISPA = DP(12) + SOP(12)*(VOLRED/100.)
SOP(11) = SOLIDPA*TSSLA
DP(11) = DISPA
SOP(13)=Q(12)*SOP(12)*(1.-TSSLA+VOLRED*TSSLA/100.-VOLRED/100.)/Q(13
)
DP(13) = DISPA
SOC(11) = VSS(11)/2.
DOC(11) = DOC(12)
SOC(13) = VSS(13)/2.
SNBC(13) = SNBC(12)*Q(12)/Q(13)
DNBC(13) = DNBC(12)
SFM(11) = SFM(12)*TSSLA
SFM(13) = (SFM(12)*(Q(13)+Q(11))-SFM(11)*Q(11))/Q(13)
DFM(11) = DFM(12)
DFM(13) = DFM(12)
VOLDIG = DIGT*Q(13) + Q(11)
AIRCFD = AIRCFD + VOLDIG*20.*60.*24.*1000./7.48
BSIZE = AIRCFD/(24.*60.)
IF(VNIT-VAER)60,60,61
60 ALK(5) =ALK(1) + 3.57*(DN(5) - DN(2))
GO TO 62
61 ALK(5) = ALK(1)
62 ALK(12) = ALK(5)
ALK(13) = ALK(12) + (DN(13) - DN(12))*3.57
ALK(15) = ALK(13)
TSS(15) = TSS(13)
VSS(15) = VSS(13)
Q(15) = Q(13)
SOC(15) = SOC(13)
SNBC(15) = SNBC(13)
SON(15) = SON(13)
SOP(15) = SOP(13)
DOC(15) = DOC(13)
DNBC(15) = DNBC(13)
DN(15) = DN(13)
DP(15) = DP(13)
SFM(15) = SFM(13)
DFM(15) = DFM(13)
RETURN
END
```

```
SUBROUTINE DRY
COMMON    Q(20),SOC(20),SNBC(20),SON(20),SOP(20),SFM(20),DOC(20),
CDNBC(20),DN(20),DP(20),DFM(20),SBOD(20),DBOD(20),CCOST(15),
CCOSTO(15),ACOST(15),AOCOST(15),NFORK(10),ASS(12),DEMBOD(10),
CVSS(20),TSS(20),ALK(20),CPERKG(15),ECF(15),Q20(10),FRPSIN(10)
COMMON DEGC,URPS,URSS,XRSS,CAER20,AEFF20,DO,CKWH,CCI,AF,GSS,WP,
CTRR,GTH,GSTH,ERR,WRE,GE,GES,TDIG,TD,VFL,TVF,SBL,NAS,NBOD,NQ,NFR
COMMON AIRCFD,FECL3,CFECL3,XMLASS,XMLBSS,XLNBSS,XMLDSS,XMLISS,VNIT
COMMON RETURN,FRDIG,C1DIG,C2DIG,CH4CFD,CO2CFD,FOOD,CAER,AEFF,CNIT,
CCEDR,CFPGAL,K35,LOOPS,FRPS,XMLSS,BOD5,VAER,BSIZE,AVF
 COMMON VDIG,ATHM,AE,VOLDIG,ARATE,AYEARS,SOLIDS,TSSLA,DIGT
COMMON XLANDC,XMU,HC,SIGMA,HO,SAND,DRYDAY,SCI,STO,STF,RC,AREA
 REAL NOTONS
 SO = TSS(13)/1000000.
 VOLUME = Q(13)*1000000.
TOTC = 1.E12
 34   CCOST1 = TOTC
HI = (HO-SAND)*SO/STO
H5 = HO-SAND
H = HI + SAND
 40   FORMAT(E6.1)
 Q(16) =Q(13)*(1. -SO/STO)
SI = SO*100.
Z =(XMU*SI*RC)/(100.*(HC**SIGMA)*SIGMA*(SIGMA+1.))
Z2 = (SIGMA + 1.)*HO*H**(SIGMA)
T =(Z*(HO**(SIGMA+1.)+SIGMA*H**(SIGMA+1.)-Z2))/3600.
WTS = 10. *(HO-SAND)*SO
UC = 500.*((SCI*WTS)**.5)
UO = 100.*((1.-STO)/STO)
UT = 100.*((1.-STF)/STF)
TDRY = (WTS/(100.*SCI))*(UO-UC+UC*LOG(UC/UT))
PREPT = 2.
TOTALT = T/24. + TDRY/24. + PREPT
RUNS = DRYDAY/TOTALT
AREA =(VOLUME/(RUNS*(HO-SAND)))*365.*2.54*12./(7.5*43560.)
NOTONS = (SO*AREA*(HO-SAND)*43560./(12.*2.54))*62.4/2000.
 CCOST(13) = 0.50*AREA*43560. + AREA*XLANDC
 CCOST(13) = CCOST(13)*CCI
CCOSTT = CCOST(13)/(NOTONS*RUNS)
 ACOST(13) = CCOST(13)*AF
 COSTO(13) =(AREA*70. +NOTONS*3.)*RUNS
COSTX = COSTO(13)/(RUNS*NOTONS)
TOTC = COSTX  + ( ACOST(13))/(NOTONS*RUNS)
 320   FORMAT(1X,3(F7.2,2X),1X,F8.2,1X,2(F7.2,3X)F9.2)
 IF(TOTC .GE. CCOST1) RETURN
HO = HO + 1.0
 Q(16) = Q(15)*(1. -SO/STO)
 TEMP1 = 0.01
```

```
5630↑ TSS(16) = TEMP1*TSS(15)
5640↑ SOC(16) = TEMP1*SOC(15)
5650↑ SNBC(16) = TEMP1*SNBC(15)
5660↑ SON(16) = TEMP1*SON(15)
5670↑ SOP(16) = TEMP1*SOP(15)
5680↑ SFM(16) = TEMP1*SFM(15)
5690↑ DOC(16) = DOC(15)
5700↑ DNBC(16) = DNBC(15)
5710↑ DN(16) = DN(15)
5729↑ DP(16) = DP(15)
5730↑ DFM(16) = DFM(15)
5780↑ GO TO 34
5800 END
6000↑END PROG
```

```
6010 PRELIMIN ARY.TRT. PRIMARY. SETTLER. AERATION TANKS...
6020 AIR.BLOW ERS.....
6030 FINAL.SE TTLER... SLUDGE.P UMPS.... CONTROL. HOUSE...
6040 SLUDGE.T HICKENER ANAEROB. DIGESTER ELUTRIAT ION.....
6050 VACUUM.. FILTER.. SLUDGE.I NCINERAT SLUDGE.D RYING.BD
6060 CHLORINA TION.... SITE.DEV ELOPMENT
6070 AEROBIC.DIGESTER
6080 0,0,0,0,0,0,0,0,0,0,0,0,0,0
6090 1.,2.,1.2,1.5,2.0,2.0,1.0,1.5,2.0,1.5,1.0,1.0,1.0,1.0,1.0
6100 1,1,0,0,0,0,0,0,0,0,0,0,0,0,0,0
6110 105.,30.,10.,2.,30.,43.,11.,19.,4.,500.,20.,250.,
6120 400.,6.0,0.02,1.0,0.05,1.0,0.01
6130 1.57,2000.,.95,60000.,750.
6140 9.0,0.76,60000.,3.0,800.,9.0,33.
6150 15.,200.,4.9,0.238,4.4
6160 25.,0.045
6170 1
6180 3000.
6190   1
6200    13.00
6210 1
6220 10.
6230   1
6240       .50
7010 0.00919,819.,.97,46.,45.,365.,10000.
7020 0.05,0.25,0.5,1.804E+9
```

APPENDIX IV

Sample Specific Resistance Data for Aerobic Sludge

TABLE A7

Typical Aerobic Sludge Buchner Funnel Data

Time (sec)	Buret Reading (ml)	Cumulative filtrate volume (ml)
0	250	0
30	233	17
60	225	25
90	220	30
120	216	34
150	213	37
180	209	41
210	206	44
240	203	47
270	200	50
300	197	53
330	194	56
360	191	59
390	189	61
420	187	63
450	185	65
480	183	67
510	180	70
540	178	72
570	176	74
600	174	76
630	172	78
660	170	80
720	166	84
780	162	88
840	159	91
900	155	95
960	152	98
1020	149	101
1080	146	104
1140	143	107
1200	140	110
1260	137	113
1320	134	116
1380	131	119
1440	129	121

TABLE A7, Continued

Time (sec)	Buret Reading (ml)	Cumulative Filtrate Volume (ml)
1500	126	124
1560	123	127
1620	121	129
1680	119	131
1740	116	134
1800	113	137
1860	112	138
1920	110	140
1980	108	142
2040	105	145
2100	103	147
2160	101	149

Temperature = 23°C

Vacuum = 10.0 cm Hg.

Filter Paper: #5 Whatman

Volume of sample = 250 ml

Linear curve fit through the data points yielded the following equation: $Y = A + BX$

where:

$A = 0.932159$

$B = 0.0905648$

$R = 0.999569$ = Coefficient of correlation

A computer program was used to fit the linear curve through the data, and another program was used to determine the specific resistance of the sludge by the following formula:

$$R = \frac{2bPA^2}{\mu c}$$

where: R = specific resistance sec^2/g

b = slope of t/V vs V curve

P = vacuum applied

A = filtration area

μ = filtrate viscosity

c = weight of solids/unit volume of filtrate

From the above equation, the value for the specific resistance was found to be 0.30698600×10^9 at 10 cm Hg.

The coefficient of compressibility is found by finding the slope of the line through values for the specific resistance at several different pressures. Another computer routine was used to determine the coefficient of compressibility after several values for the specific resistance were found. The value obtained for the coefficient of compressibility was: 0.97121706, obtained from four different pressure experiments.

APPENDIX V

Treatment Plant Cost Curves

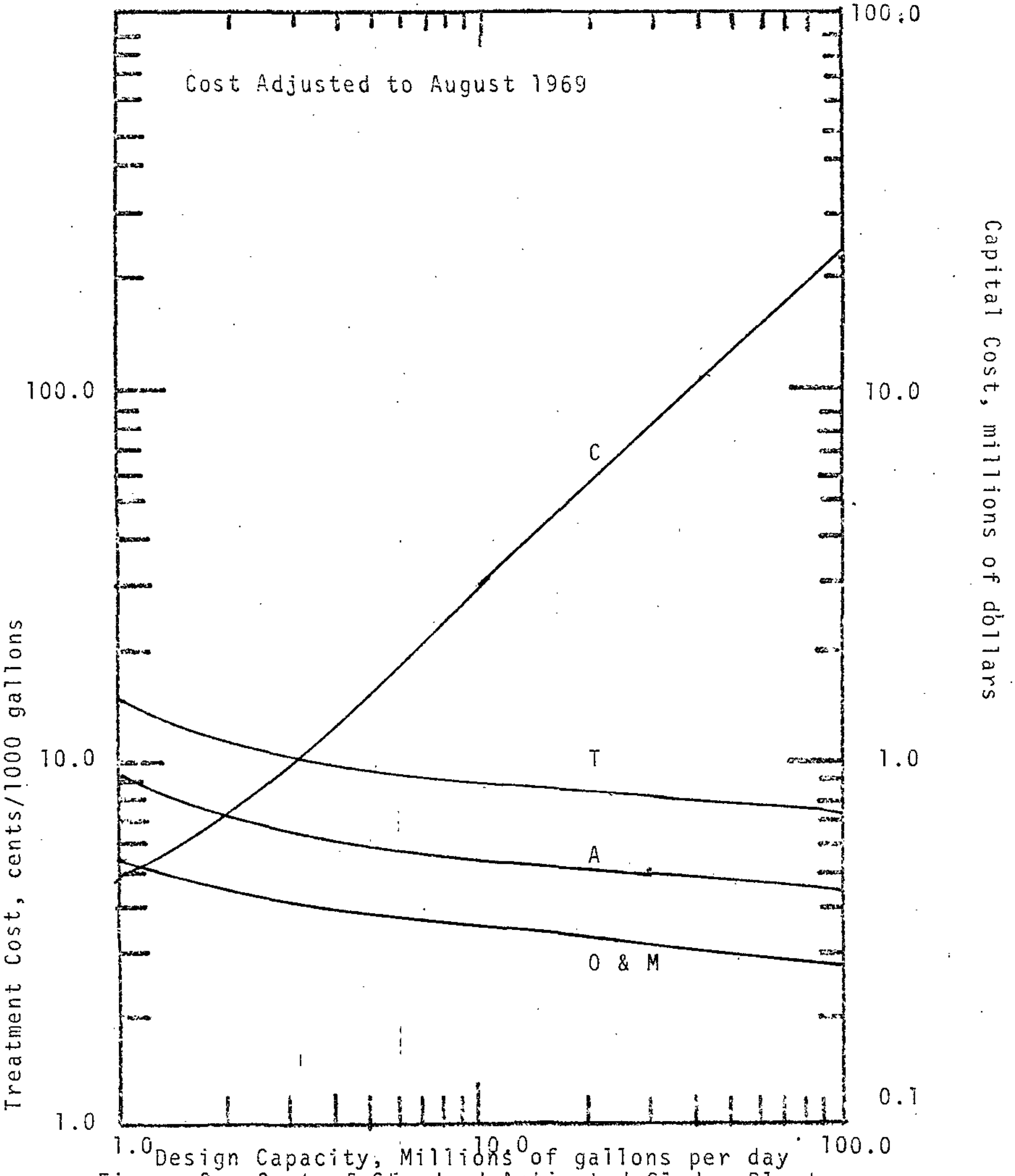

Figure 3. Cost of Standard Activated Sludge Plant

C = Capital Cost, millions of dollars.
A = Debt Service, cents per 1000 gallons.
O&M = Operating andaMaintenace Cost, cents per 1000 gallons.
T = Total Treatment Cost, cents per 1000 gallons

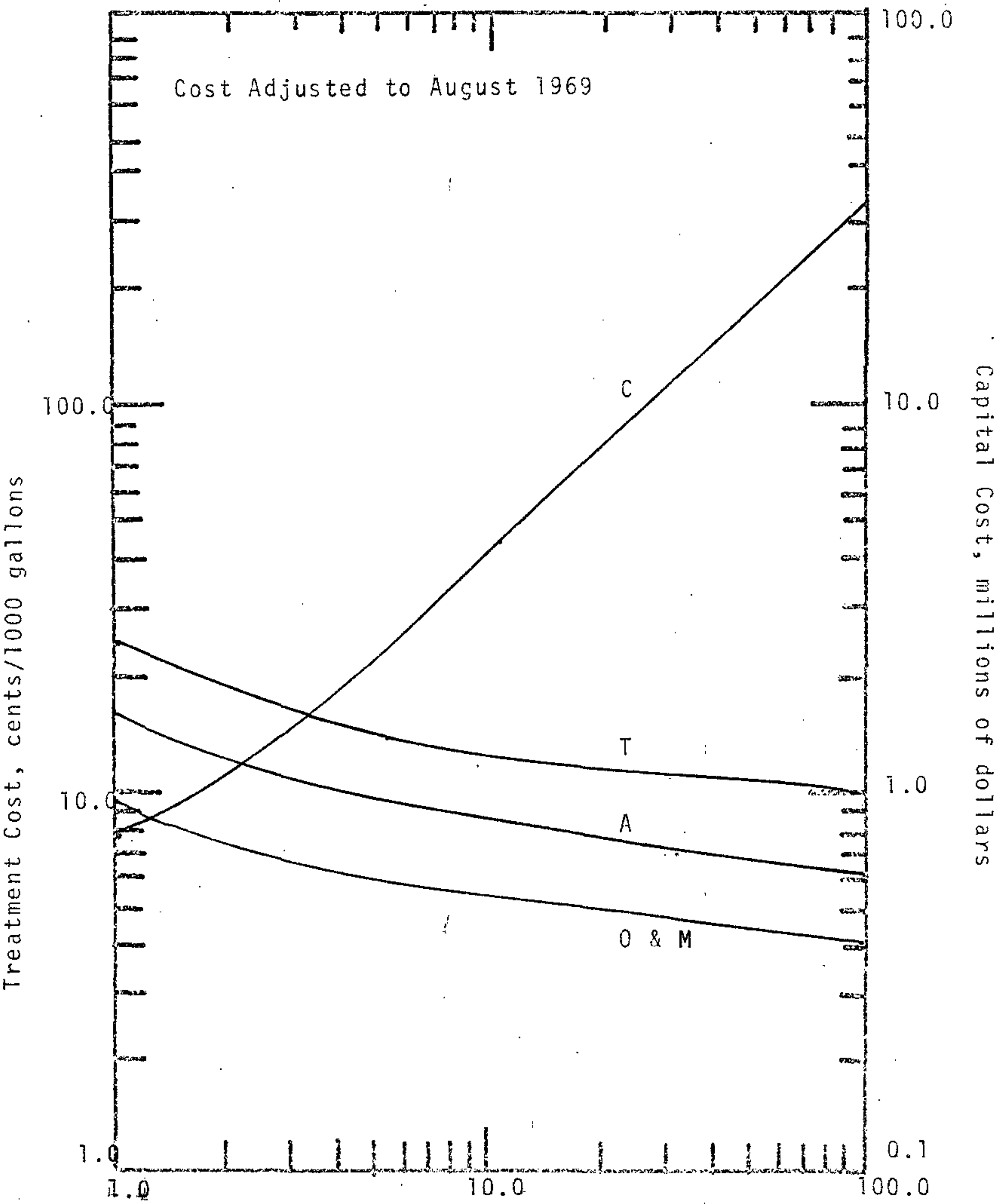

Figure 4. Cost of Activated Sludge Plants without Primary Treatment.

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

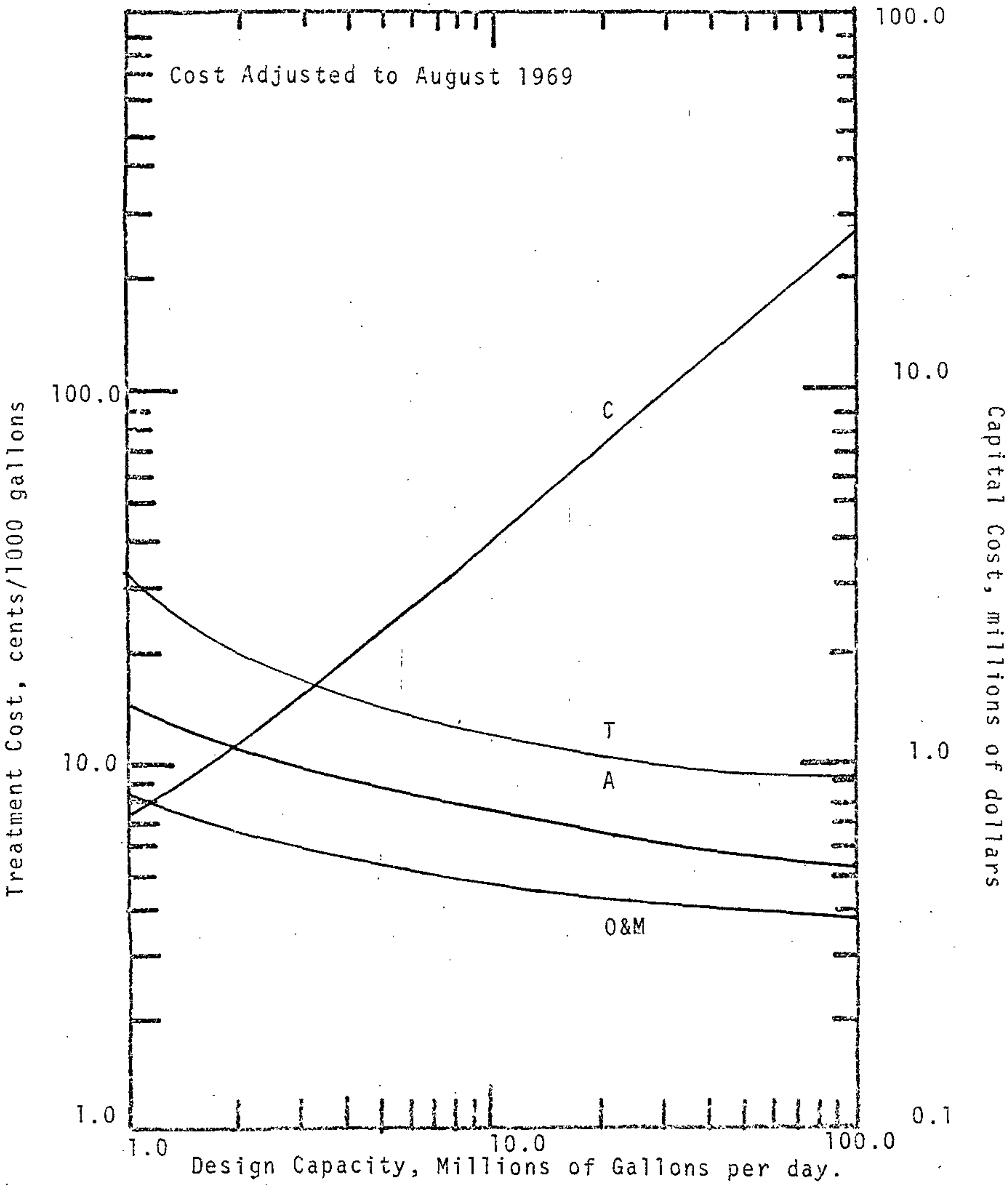

Figure 5. Cost of Activated Sludge Plants Utilizing Aerobic Digestion

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

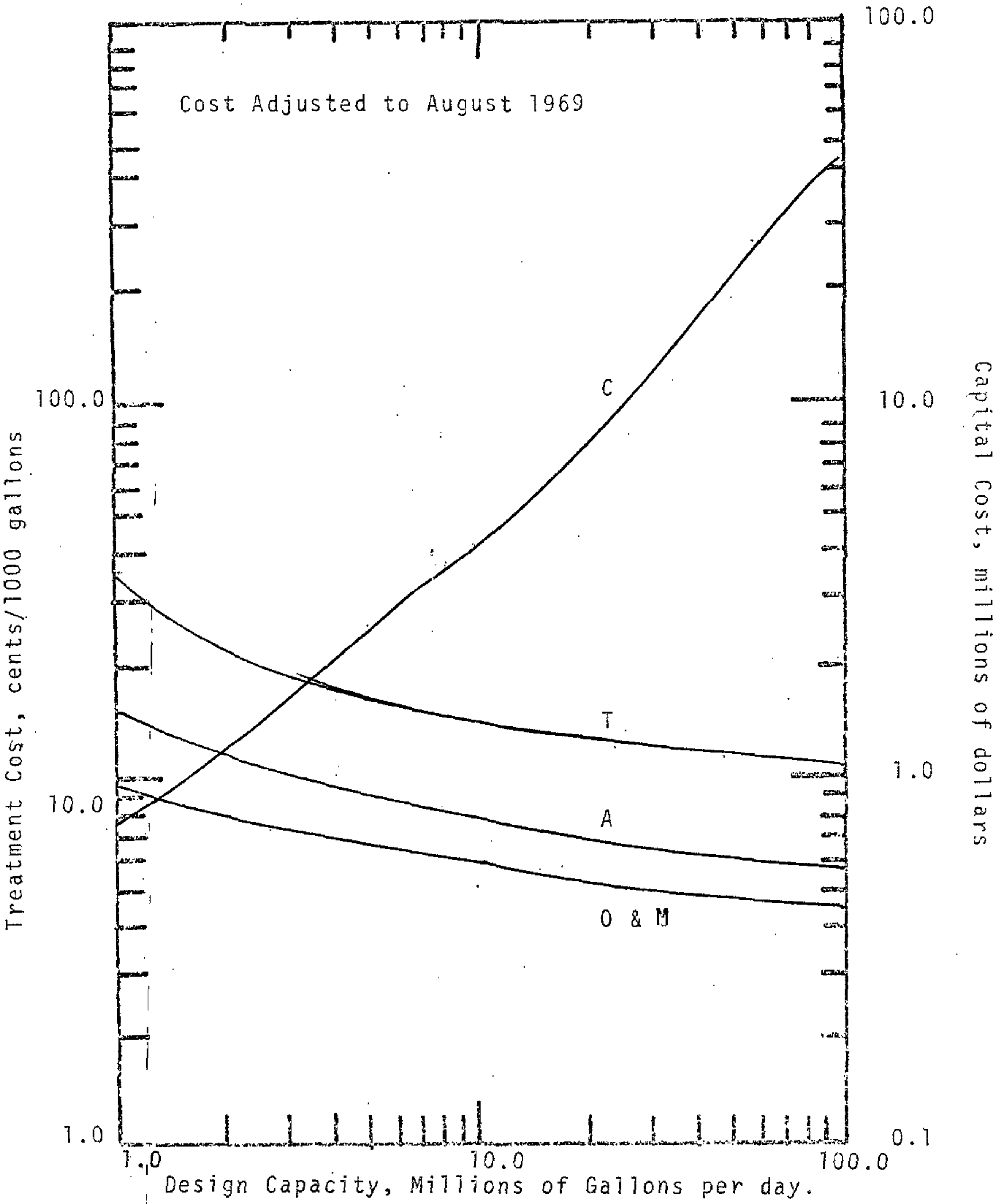

Figure 6. Cost of Activated Sludge Plants Utilizing Anaerobic Digestion and Sand Sludge Beds

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

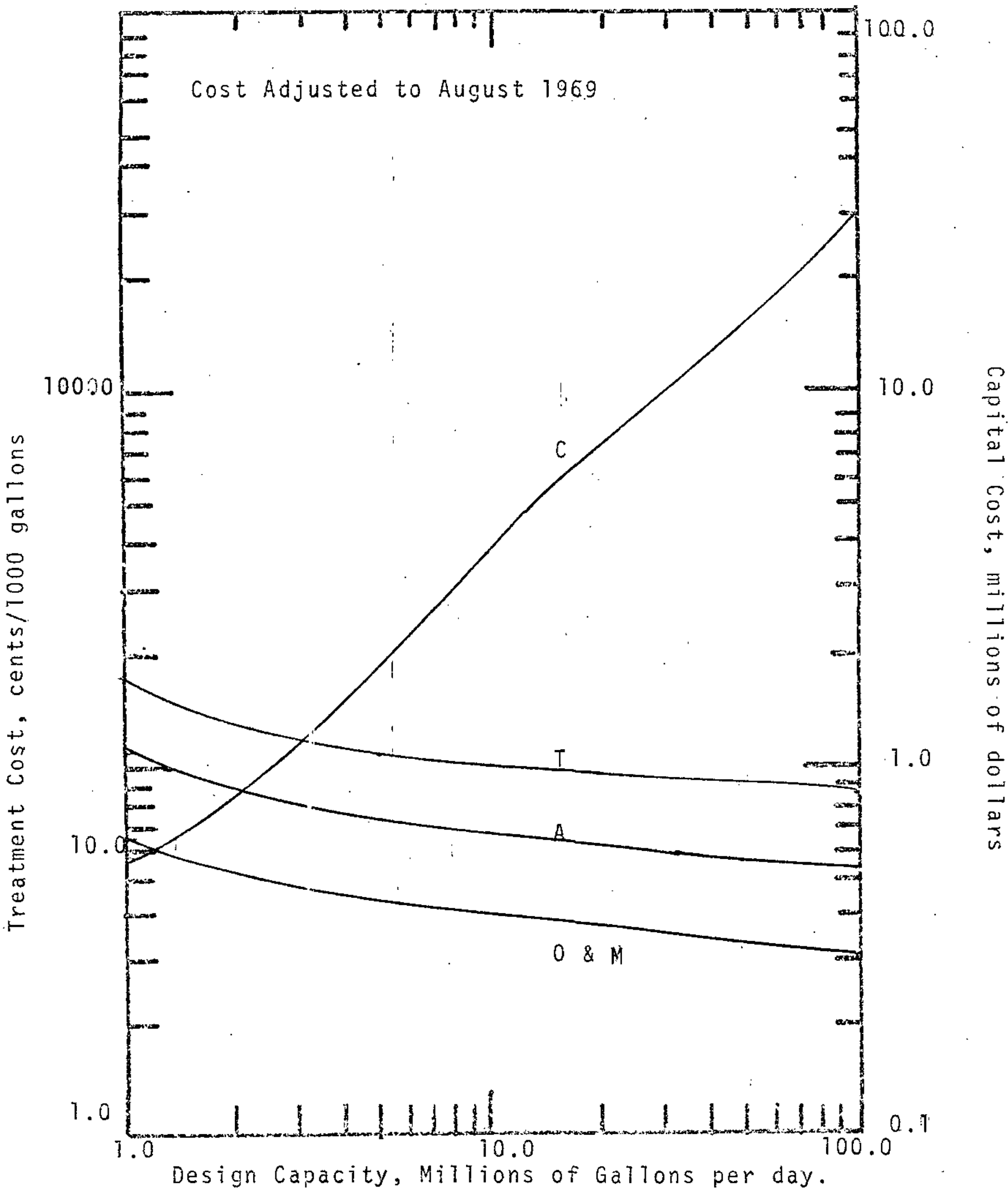

Figure 7. Cost of Activated Sludge Plants Omitting Primary Sedimentation, and Using Aerobic Digestion and Sand Sludge Beds.

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

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