JOURNAL OF THE ENVIRONMENTAL ENGINEERING DIVISION

Cost Interactions in Activated Sludge Systems

By Michael K. Stenstrom,[1] A. M. ASCE and John F. Andrews,[2] M. ASCE

Introduction

Wastewater treatment plants are complicated systems consisting of several interacting processes. The design, operation, and control of these plants is not a simple task, since most of the processes that make up the plant, and the interactions among these processes, are not precisely understood. Moreover, the environment in which the plant functions is continually changing, and is often uncertain. However, the analysis of complex systems, such as a wastewater treatment plant, has been facilitated by new analytical techniques, and computers.

Several investigators have focused their attention on the problems of optimal plant design and operation. Smith (16) was one of the earliest to study the effects of interactions in processes. He developed a steady-state design program that was capable of sizing the unit processes normally found in wastewater treatment plants. His program also estimated the cost of the facilities, given baseline cost parameters. Parker and Dague (12) developed an optimal, steady-state design procedure for wastewater treatment plants composed of a primary clarifier, an activated sludge process, and an aerobic digester. These authors were able to show the effects of the design mixed-liquor suspended solids (MLSS) concentration upon the total system cost. Their results indicate that, before the thickening criteria of the secondary clarifier becomes limiting, total capitalized cost tends to decrease as MLSS concentration increases.

Middleton and Lawrence (8,9) have also analyzed treatment plant design, using steady-state models with a numerical search procedure to determine optimal, least-cost design. They performed a sensitivity analysis to determine the effects of operating modes and input parameters, such as mean cell retention time, wastewater chemical oxygen demand, and methods of sludge disposal. Their

Note.—Discussion open until January 1, 1981. To extend the closing date one month, a written request must be filed with the Manager of Technical and Professional Publications, ASCE. This paper is part of the copyrighted Journal of the Environmental Engineering Division, Proceedings of the American Society of Civil Engineers, Vol. 106, No. EE4, August, 1980. Manuscript was submitted for review for possible publication on October 16, 1979.

[1]Asst. Prof., Engrg. Systems Dept., Univ. of California, Los Angeles, Los Angeles, Calif.

[2]Prof., Dept. of Civ. Engrg., Univ. of Houston, Houston, Tex.

analysis indicates that total system cost generally increases with increasing mean cell retention time, and that the trend is very sensitive to certain key parameters, such as sludge disposal costs.

Mynhier and Grady (10) have developed plant design techniques using nomographs which were based upon steady-state models. Tarrer, et al. (20) have developed similar design procedures, and Grady (5) has illustrated the technique of dynamic programming to calculate optimal plant designs for economic input parameters. Kincannon, et al. (6) have investigated optimal plant design from an energy basis.

The previously cited investigators all addressed the problems of plant design and operation, with economic input parameters influencing treatment system design and operation. However, many treatment plants have already been constructed, and plant operation is constrained within the limits set by design, such as recycle sludge thickening ability, or oxygen transfer capability.

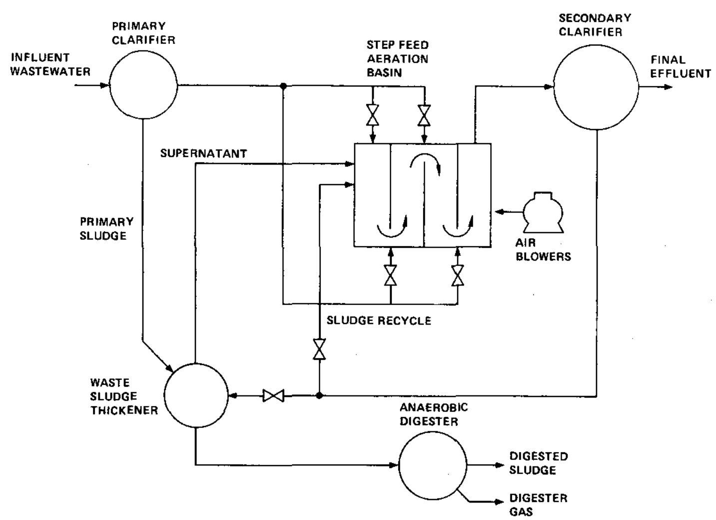

FIG. 1.—Wastewater Treatment Plant Flow Chart

The results reported here are only a portion of a larger investigation (17) directed at the development of real-time process control strategies. During the course of this investigation, the interactions among three major economic variables, i.e., oxygen transfer cost, anaerobically digested sludge disposal cost, and methane digestion gas value, were also determined.

The analyses reported herein are the results of a mathematical model, and simulation of a typical wastewater treatment plant as shown in Fig. 1. Dynamic models were used for each process, and real-time control techniques were developed. Finally, the process control techniques were integrated into an overall plant control strategy. The previously cited operating variables were calculated and integrated over several weeks of operation at mean cell retention times, between five and ten days. The cost of oxygen transfer and sludge disposal, and the value of methane gas at each value of mean cell retention time, were then compared.

This investigation differs from previous investigations in that only the effects of changing plant operations were determined. The range of plant operation was constrained within fixed limits, since process sizes were not changed. The size of the secondary clarifier limited process operation to a maximum of 10 days.

Dynamic Model Description

The dynamic model for the plant shown in Fig. 1 has been described previously (16,19), therefore only a brief description will be given here. The plant sizes and dimensions were as follows:

1. Primary clarifier with an area = 8,000 ft^2 (743 m^2), a depth = 12 ft (3.66 m), and an overflow rate = 1,250 gal/ft^2 day (51 m^3/m^2 day) (average).
2. Aeration basin with a volume = 3.6×10^6 gal (1.36×10^4 m^3) and a hydraulic retention time = 9 h (average time).
3. Secondary clarifier with an area = 12,500 ft^2 (1,161 m^2), a depth = 12 ft (3.66 m), and an overflow rate = 800 gal/ft^2 day (34 m^3/m^2 day).
4. Anaerobic digester with a volume = 2.6×10^6 gal (9.84×10^3 m^3).
5. Waste sludge thickener with a surface area = 5,000 ft^2 (465 m^2).
6. Input parameters consist of (values are averages): Flow = 10 mgd (0.44 m^3/s); BOD_u = 250 mg/l; BOD_5 = 170 mg/l; TSS = 150 mg/l; and NH_3-N = 30 mg/l.

The inputs to the model were described by Fourier time-series models based upon data collected from the cities of Atlanta and Minneapolis-St. Paul. Influent

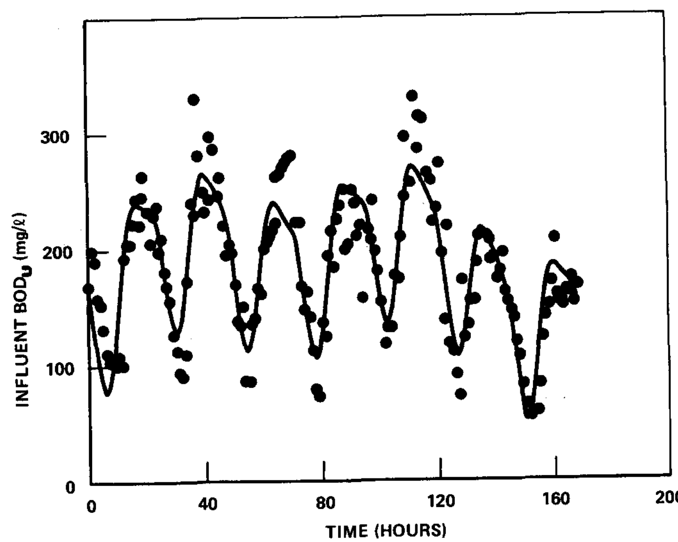

FIG. 2.—Influent Biochemical Oxygen Demand

flow rate, BOD_5, suspended solids, and ammonia nitrogen were described. Fig. 2 shows the time series model and the data for the BOD_5 input function, and is typical of the other time series models.

The primary clarifier model, develped by Bryant (2), was used in this investigation. The suspended solids removal rate was predicted by an empirical expression developed from the data presented by Thereoux and Betz (21). The hydraulics were simulated by a tanks-in-series model and the underflow solids concentration was assumed to be constant at 5%.

The aeration basin was modeled as four completely mixed tanks in series, with a structural model for the microbial mass, similar to the technique of Busby and Andrews (3). A nitrification model, based upon the work of Poduska and Andrews (14), was also incorporated into the aeration basin model. A material balance on dissolved oxygen was included in the aeration basin model in order to simulate dissolved oxygen control, and specific oxygen uptake rate control

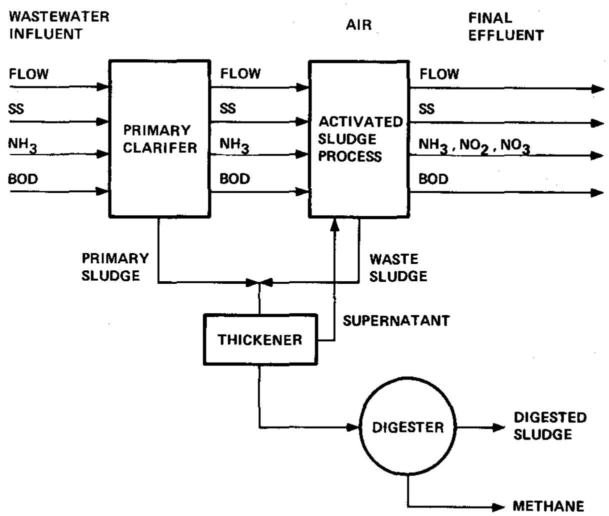

FIG. 3.—Overview of Plant Models and Information Flow

(18). The secondary clarifier was modeled in a fashion similar to the technique developed by Bryant (2). Effluent suspended solids were calculated using an empirical relationship based upon the data of Pflanz (3).

The sludge thickener model was developed from the previous work of Smith (16). The anaerobic digester model was developed using the kinetic relationships, and stoichiometry, employed by McCarty and his co-workers (7,11). Using these relationships, it was possible to estimate the volatile solids destruction and methane gas production. The information flow for the entire model is shown in Fig. 3.

Plant Control Strategy

The treatment plant was simulated, and control strategies were developed to control the activated sludge process at a constant growth rate (17–19). Once

suitable control strategies were developed, the model was used to determine the weekly average values of digested sludge production, methane gas generation, and oxygen requirements. Mean cell retention times, from five to ten days, were simulated. Ten days was the maximum mean cell retention time which could be maintained without failure of the secondary clarifier. The effluent BOD_5 concentration was approximately the same for each mean cell retention time, between five and ten days. The model predicted a decrease in soluble effluent substrate with increasing mean cell retention time, but effluent suspended solids, which were assumed to contribute to effluent BOD_5, increased with increasing mean cell retention time. This is a direct result of using the Pflanz relationship for calculating effluent suspended solids. The weekly averages of

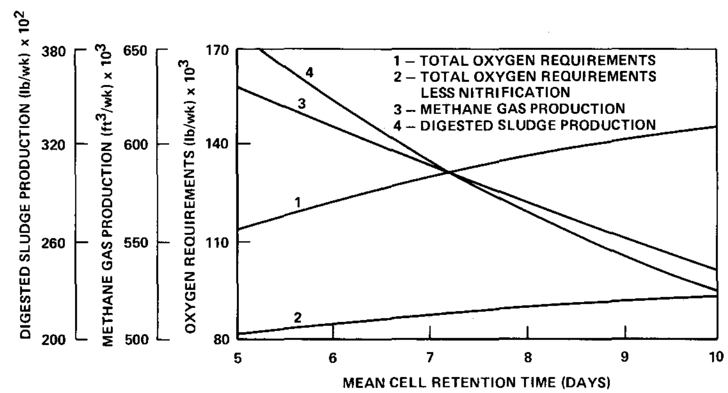

FIG. 4.—Operating Conditions as Function of Mean Cell Retention Time

oxygen requirements, sludge production, and methane gas generation are shown in Fig. 4.

Weekly Cost Comparisons

The total operating cost of a plant depends upon many factors, including labor, the cost of power for pumping and aeration, the value of the methane gas produced, the cost of ultimate sludge disposal, and other factors. Using the results of the plant simulations shown in Fig. 4, it is possible to evaluate a partial operating cost for different base values of sludge disposal cost, oxygen transfer, and the value of methane gas. These three factors often represent a large portion of total operating costs. Figs. 5–8 are contour plots that show the weekly operating costs for the varying costs and values assigned to the three previously mentioned variables. In calculating the costs for Figs. 5–8, it was assumed that the costs associated with these three variables are controllable. Many plants have the ability to control these costs, because they have turn-down or turn-up capacity on the aeration systems, facilities to use excess digester gas, or controllable expenses associated with sludge disposal, such as chemical costs or trucking expenses. Many other plants do not have this ability; for these plants, the results shown in Figs. 5–8 may not be applicable.

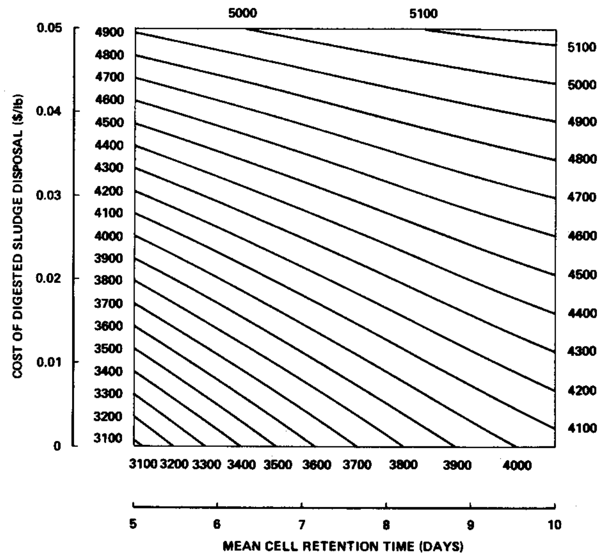

FIG. 5.—Weekly Cost with Changing Sludge Disposal Cost as Function of Mean Cell Retention Time (Numbers at Graph Edges Denote Contour Values, in dollars per week)

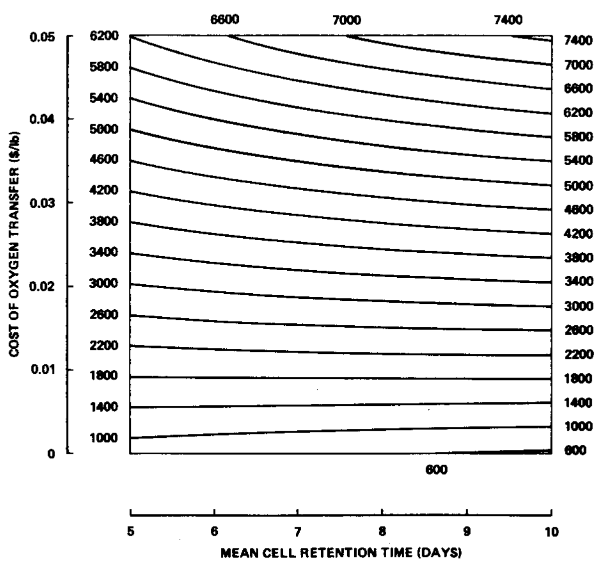

FIG. 6.—Weekly Cost with Changing Oxygen Transfer Cost as Function of Mean Cell Retention Time (Numbers at Graph Edges Denote Contour Values, in dollars per week)

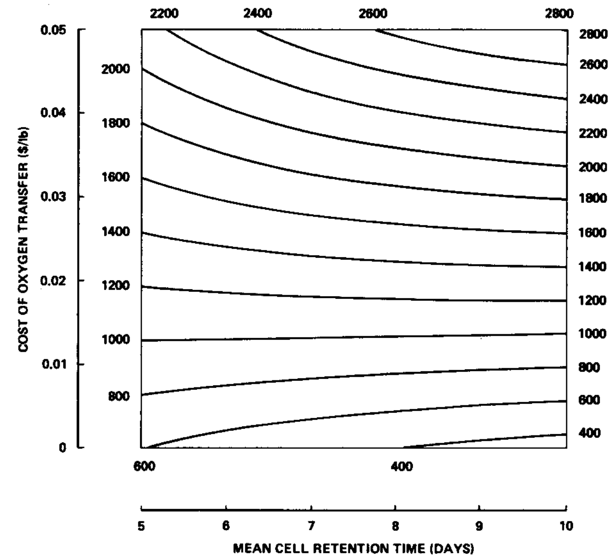

FIG. 7.—Weekly Cost with Changing Oxygen Transfer Cost as Function of Mean Cell Retention Time [Without Nitrification (Numbers at Graph Edges Denote Contour Values, in dollars per week)]

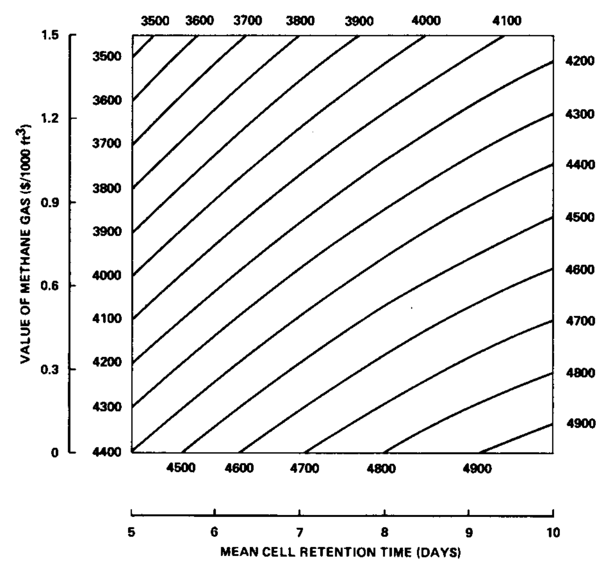

FIG. 8.—Weekly Cost with Changing Digester Gas Value as Function of Mean Cell Retention Time (Numbers at Graph Edges Denote Contour Values, in dollars per week)

Fig. 5 shows the weekly costs as a function of sludge age and sludge disposal costs. For this figure, a value of \$0.72/1,000 ft^3 at standard temperature and pressure (STP) as CH_4 (\$25.4/1,000 m^3) was assigned as the value of the digester gas, and the cost for oxygen transfer was assumed to be \$0.030/lb O_2 (\$0.066/kg O_2). Ultimate sludge disposal cost ranged from \$0/dry ton to \$100/dry ton (\$0/kg to \$0.11/kg). This range represents the probable range of several sludge treatment and disposal alternatives (15). The results show that, at low sludge disposal costs, there is approx \$1,000/week savings resulting from operation at low mean cell retention time. At higher sludge disposal costs, the differences in economics of low versus high mean cell retention time become less pronounced. Although not shown on this contour figure, operation at high mean cell retention time becomes more economic than low mean cell retention time operation, when oxygen transfer costs are less.

Fig. 6 shows the effect of changing oxygen transfer costs in process operation. This figure was calculated using a sludge disposal cost of \$50/dry ton (\$0.055/kg), and a methane gas value of \$0.72/1,000 ft^3 (\$25.4/1,000 m^3). At low values of oxygen transfer cost, high mean cell retention time operation is preferred; however, at high oxygen transfer costs, low mean cell retention time is preferred, resulting in over a \$1,000/week cost difference. In the plant simulations it was assumed that the temperature was moderatly high (20° C); therefore nitrification occurred even at five days mean cell retention time. However, in colder climates nitrification might not occur. Fig. 7 shows this situation, with all cost parameters the same. This figure shows that the increase in cost for nitrification can be very significant.

Fig. 8 shows the effect of changing digester gas value on operating cost. For this figure the sludge disposal cost was \$50/dry ton (\$0.055/kg), and the oxygen transfer cost was \$0.03/lb O_2 (\$0.066/kg O_2). Digester gas values ranged from zero to \$1.50/1,000 ft^3 (\$52.9/1,000 m^3). The increased value of methane gas can result in a weekly cost difference of approx \$1,000. The cost difference of low versus high mean cell retention time operation increases about \$200/week with increasing digester gas value. Recent work by Brown and Caldwell (1) has indicated that methane can be produced from digestion of classified municipal refuse for \$2.5 to \$5.0/1,000 ft^3 (\$88.3 to \$177/1,000 m^3). Production of gas at this price is considered economical in areas where the gas can be used for augmenting other fuels, such as oil and coal, for electric power production. This is especially true in areas such as the Los Angeles basin, where air quality problems exist. The writers believe that the value of digester gas will increase rapidly to the values indicated in Fig. 8, and perhaps to even higher values in the near future.

Conclusions

The previous figures and analysis have shown that there can be an economical mean cell retention time that can result in significant operational cost savings. In the hypothetical cases, it was shown that the least cost operating strategy changed for the different, but probable, values of the digester gas value, oxygen transfer cost, and the ultimate sludge disposal cost. Moreover, the change in plant operation required to achieve savings is within the operating range of many, if not most, treatment systems. For these treatment systems, few if

any design changes may be required to achieve more economical operation.

The range of operational possibilities that exists is far more complex and difficult to analyze than this paper might make it seem. Although an advanced, dynamic mathematical model, which considers such effects as stochastic inputs, changing sludge concentration, varying oxygen transfer efficiency, etc., was used, the analysis presented here is far from comprehensive. There are many other factors, such as the overdesign of certain operations in the treatment system, in anticipation of expansion, which have not been explored. It appears that there may be even greater benefits if a case-by-case analysis is performed.

A second logical conclusion from this research is that there is a need for operation flexibility in treatment systems. The economic incentives which dictate system operation may change rapidly, perhaps even seasonally. A treatment system design which allows operational flexibility will enable operators to operate their plants more economically, under changing and unpredictable financial environments. The capital cost of new facilities and modifications, required to permit operation flexibility, should be balanced with expected savings.

Acknowledgments

The writers are indebted to the National Science Foundation, whose support made this work possible.

Appendix I.—References

1. Brown, and Caldwell, "Microbial Production of Methane from Refuse," report prepared for Southern California Edison Company, Walnut Creek, Calif., July, 1978.
2. Bryant, J. O., "Continuous Time Simulation of the Conventional Activated Sludge Wastewater Renovation System," thesis presented to Clemson Univ., Clemson, S.C., in 1972, in partial fulfillment of the requirements for the Degree of Doctor of Philosophy.
3. Busby, J. B., and Andrews, J. F., "Dynamic Modeling and Control Strategies for the Activated Sludge Process," *Journal of the Water Pollution Control Federation,* Vol. 47, No. 5, May, 1975, pp. 1055–1080.
4. *Continuous System Modeling Program III, (CSMP III) Program Reference Manual,* Program No. 5734-X59, Ontario, Canada, IBM International Business Machines, 1972.
5. Grady, C. P. L., "Simplified Optimization of the Activated Sludge Process," *Journal of the Environmental Engineering Division,* ASCE, Vol. 103, No. EE3, Proc. Paper 12974, June, 1977, pp. 413–429.
6. Kincannon, D. F., Gaudy, A. F., De Fries, R. N., and Headstream, M., "Operation of Treatment Processes—Energy Consideration," *Energy Conservation in the Design of Water Quality Control Facilities,* Dept. of C.E., University of Missouri, Columbia, Mo., 1976, pp. 120–126.
7. McCarty, P. L., "Anaerobic Waste Treatment Fundamentals, Part Four—Process Design," *Public Works,* Vol. 99, No. 12, Dec., 1964, pp. 95–99.
8. Middleton, A. C., and Lawrence, A. W., "Cost Optimization of Activated Sludge Systems," *Biotechnology and Bioengineering,* Vol. 16, No. 6, June, 1974, pp. 807–826.
9. Middleton, A. C., and Lawrence, A. W., "Least Cost Design of Activated Sludge System," *Journal Water Pollution Control Federation,* Vol. 48, No. 5, May, 1976, pp. 889–905.
10. Mynhier, M. D., and Grady, C. P. L., "Design Graphs for Activated Sludge Process," *Journal of the Environmental Engineering Division,* ASCE, Vol. 101, No. EE5, Proc. Paper 11659, Oct., 1975, pp. 829–846.
11. O'Rourke, J. T., and McCarty, P. L., "Anaerobic Treatment at Reduced Temperatures," paper presented at the October, 1974, Water Pollution Control Federation Meeting.

12. Parker, G. F., and Dague, R. R., "Optimal Design of Wastewater Treatment Systems by Enumeration," *Journal of the Sanitary Engineering Division,* ASCE, Vol. 98, No. SA6, Proc. Paper 9403, Dec., 1972, pp. 833–851.
13. Pflanz, P., "Performance of (Activated Sludge) Secondary Sedimentation Basins," *Advances in Water Pollution Research,* Vol. 4, S. H., Jenkins, ed., Pergamon Press, New York, N.Y., 1969, pp. 569–581.
14. Poduska, R. A., and Andrews, J. F., "Dynamics of Nitrification in the Activated Sludge Process," *Journal of the Water Pollution Control Federation,* Vol. 47, No. 11, Nov., 1975, pp. 2500–2619.
15. "Sludge Management Alternatives for the Los Angeles/Orange County Metropolitan Area," LA/OMA Project, Whittier, Calif., May, 1977.
16. Smith, R., "Preliminary Design of Wastewater Treatment Systems," *Journal of the Sanitary Engineering Division,* Vol. 95, No. SA1, Proc. Paper 6426, Feb., 1969, pp. 117–145.
17. Stenstrom, M. K., "A Dynamic Model and Computer Compatible Control Strategies for Wastewater Treatment Plants," thesis presented to Clemson University, Clemson, S.C., 1976, in partial fulfillment of the requirements for the Degree of Doctor of Philosophy.
18. Stenstrom, M. K., "Real-Time Control of the Activated Sludge Process," *Journal of the Environmental Engineering Division,* ASCE, Vol. 105, No. EE2, Proc. Paper 14521, Apr., 1979, pp. 245–260.
19. Stenstrom, M. K., and Andrews, J. F., "Dynamic Models and Control Strategies for Wastewater Treatment Plants—Recent Advances," presented at the August, 1977, IFAC meeting, held at Kyoto, Japan.
20. Tarrer, A. R., Grady, C. P. L., Lim, H. C., and Koppel, K. B., "Optimal Activated Sludge Design under Uncertainty," *Journal of the Environmental Engineering Division,* ASCE, Vol. 102, No. EE3, Proc. Paper 12201, June, 1976, pp. 657–673.
21. Thereoux, R. J., and Belz, J. M., "Sedimentation and Preaeration at Los Angeles," *Sewage and Industrial Wastes,* Vol. 31, No. 11, Nov., 1959, pp. 1259–1266.

Appendix II.—Notation

The following symbols are used in this paper:

A = pounds of volatile suspended solids produced per day in digester;
C = suspended solids concentration in Secondary Changer;
e = efficiency of waste utilization in digester;
F = ultimate BOD added per day to digester;
RC = volume of methane gas produced per day;
S = substrate;
t = time;
V_s = suspended solids settling velocity;
u = fluid velocity, relative to clarifier walls;
XA = active mass concentration;
XI = inert mass concentration;
XS = stored mass concentration; and
Z = (verticle) distance.