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AN INTEGRATED CONTROL STRATEGY FOR THE ACTIVATED SLUDGE PROCESS

BY

ZDENKO ZDENKO VITASOVIC

A THESIS SUBMITTED
IN PARTIAL FULFILLMENT OF THE
REQUIREMENTS FOR THE DEGREE

DOCTOR OF PHILOSOPHY

APPROVED, THESIS COMMITTEE:

John F. Andrews, Professor,
Environmental Engineering,
Chairman

P. B. Bedient, Associate Professor,
Environmental Engineering

J. W. Hightower, Professor,
Chemical Engineering

HOUSTON, TEXAS

MAY, 1986
An Integrated Control System for the Activated Sludge Process

by

Zdenko Vitasovic

ABSTRACT

The objective of this study was to develop an integrated control strategy for the activated sludge process. For the purpose of this study, the process included the influent pump station, the biological reactor, the air supply and distribution system, and the solids-liquid separator.

A comprehensive dynamic model was developed for the process and control was superimposed on the model. The overall model consisted of deterministic and stochastic components. The models for individual process units were taken from the literature and modified where necessary to include features necessary for control strategy development. Computer simulations were used to investigate the control strategy, interactions between the control loops, interactions between process units, and between design and operation. A limited amount of data was available from full scale experimentation conducted at a wastewater treatment plant in Houston, Texas.

The proposed overall control system is modular in nature and includes modules for each unit of the process. The control
strategy has been designed for application in normal, routinely encountered conditions that prevail most of the time. The objective function was to minimize operating costs with permit limits being treated as constraints. Occasional upset conditions, usually the result of a significant disturbance to the process or equipment failure, are recognized by the control system and special procedures are activated in order to avoid gross process failure.

The control strategy for the biological reactor was based on a definition of process state consisting of a combination of DO concentration and specific oxygen uptake rate. A rule based controller was designed to maintain the optimal process state.

Computer simulations indicated that control of flow between the biological reactor and the separator was capable of eliminating high frequency variations in flow rate. This strategy was proposed for minimizing adverse effects on clarification. Based upon examination of full scale plant data, it was concluded that a strong stochastic component is present for low concentrations of effluent suspended solids. Deterministic models would therefore be inadequate for prediction of low concentrations (<10 mg/L) of solids in the effluent. A deterministic model was found to be adequate to describe thickening.

Analysis of the full-scale data shows that for low-friction-head, high-static-head pump stations, on/off pumping
requires less energy than variable speed pumping or automatic throttling of the discharge. Due to the small size of the wet well at the station studied, variable speed pumping was found to be incapable of significantly dampening variations in influent flow rate.

There are strong interactions between design and operation of treatment plants. The potential for improvement in the operation of an existing plant through implementation of process control is thus highly dependent on the plant design. Interactions between individual units of the activated sludge process and between control loops were examined and found to have a pronounced effect on the overall control system performance. It was concluded that it is important to understand these interactions and to consider them during the design of both treatment plants and control strategies.
ACKNOWLEDGMENTS

I wish to acknowledge my major advisor, Dr. John F. Andrews for his assistance during the course of this research. I also wish to express my appreciation to the other members of my committee, Professors Phil Bedient and Joe Hightower.

I wish to thank my fellow graduate students and friends for their contributions, both professional and personal.

I wish to thank the National Science Foundation and the City of Houston for financial sponsorship of this research. I especially thank Dr. M.T. Garrett of the City of Houston, Wastewater Division and Mr. Walter Schuk of the USEPA Water Engineering Research Laboratory in Cincinnati for their support and sharing of their practical knowledge.

I would like single out the invaluable assistance of Drs. Peter Dold, Larry McGaughey, Mike Stenstrom and Gustaf Olsson.
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<td>cross sectional area of the wet well (L**2)</td>
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<td>Area</td>
<td>cross sectional area of the separator (L**2)</td>
</tr>
<tr>
<td>ARFLOW</td>
<td>air flow rate (L**3/T)</td>
</tr>
<tr>
<td>B</td>
<td>flow imbalance (L**3/T)</td>
</tr>
<tr>
<td>bh</td>
<td>endogenous respiration rate</td>
</tr>
<tr>
<td>C</td>
<td>solids concentration (M/L**3)</td>
</tr>
<tr>
<td>COD</td>
<td>chemical oxygen demand (M/L**3)</td>
</tr>
<tr>
<td>D</td>
<td>dispersion coefficient (L**2/T)</td>
</tr>
<tr>
<td>DO</td>
<td>dissolved oxygen concentration (M/L**3)</td>
</tr>
<tr>
<td>DOS</td>
<td>DO saturation concentration (M/L**3)</td>
</tr>
<tr>
<td>dz</td>
<td>finite element height (L)</td>
</tr>
<tr>
<td>E</td>
<td>expectation</td>
</tr>
<tr>
<td>Ed</td>
<td>efficiency of the variable speed drive (%)</td>
</tr>
<tr>
<td>Eff</td>
<td>overall efficiency (%)</td>
</tr>
<tr>
<td>Em</td>
<td>efficiency of the motor (%)</td>
</tr>
<tr>
<td>Ep</td>
<td>pump efficiency (%)</td>
</tr>
<tr>
<td>f</td>
<td>nonbiodegradable fraction of active mass (M/M)</td>
</tr>
<tr>
<td>F</td>
<td>flow rate (L**3/T)</td>
</tr>
<tr>
<td>F0</td>
<td>influent flow rate (L**3/T)</td>
</tr>
<tr>
<td>Flow</td>
<td>flow through the pump (L**3/T)</td>
</tr>
<tr>
<td>fma</td>
<td>max. fraction of substrate in active mass (M/M)</td>
</tr>
<tr>
<td>Fout</td>
<td>effluent flow rate from the separator (L**3/T)</td>
</tr>
<tr>
<td>fp</td>
<td>stored particulate substrate fraction (M/M)</td>
</tr>
<tr>
<td>Fr</td>
<td>recycle flow rate (L**3/T)</td>
</tr>
<tr>
<td>fs</td>
<td>fraction of stored mass (M/M)</td>
</tr>
<tr>
<td>fshat</td>
<td>maximum fraction of stored mass (M/M)</td>
</tr>
<tr>
<td>ga</td>
<td>specific weight of water (M/L**3)</td>
</tr>
<tr>
<td>gam</td>
<td>autocovariance</td>
</tr>
<tr>
<td>Gs</td>
<td>settling flux (M/L**2/T)</td>
</tr>
<tr>
<td>Gup</td>
<td>upward movement of solids (M/L**2/T)</td>
</tr>
<tr>
<td>h</td>
<td>level in the wet well (L)</td>
</tr>
<tr>
<td>Ht</td>
<td>total dynamic head (Force/L**2)</td>
</tr>
<tr>
<td>k</td>
<td>lag</td>
</tr>
<tr>
<td>K</td>
<td>constant</td>
</tr>
<tr>
<td>Ka</td>
<td>particulate substrate adsorption constant (L**3/M/T)</td>
</tr>
<tr>
<td>Kfs</td>
<td>saturation coefficient</td>
</tr>
<tr>
<td>Kf</td>
<td>oxygen transfer rate coefficient (l/T)</td>
</tr>
<tr>
<td>Kmp</td>
<td>max. substrate utilization rate constant (M/M/T)</td>
</tr>
<tr>
<td>Kms</td>
<td>maximum specific growth constant (M/M/T)</td>
</tr>
<tr>
<td>Kn</td>
<td>stoichiometric coefficient</td>
</tr>
<tr>
<td>Kp</td>
<td>hydrolysis rate coefficient (l/T)</td>
</tr>
<tr>
<td>Ks</td>
<td>substrate saturation coefficient (M/L**3)</td>
</tr>
<tr>
<td>Ksp</td>
<td>saturation coefficient</td>
</tr>
<tr>
<td>Kss</td>
<td>rapidly biodegr. substr. saturation coefficient</td>
</tr>
<tr>
<td>Kt</td>
<td>transport rate coefficient (L**3/M/T)</td>
</tr>
<tr>
<td>mhnb</td>
<td>nitroreactor maximum growth rate (l/T)</td>
</tr>
<tr>
<td>mhnss</td>
<td>nitrosomonas maximum growth rate (l/T)</td>
</tr>
<tr>
<td>Min</td>
<td>minimum function</td>
</tr>
<tr>
<td>MLSS</td>
<td>mixed liquor suspended solids (M/L**3)</td>
</tr>
<tr>
<td>mu</td>
<td>mean of the stochastic process</td>
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</tbody>
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List of Symbols (cont'd)

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
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<tbody>
<tr>
<td>nh4</td>
<td>ammonia concentration (M/L**3)</td>
</tr>
<tr>
<td>no2</td>
<td>nitrite concentration (M/L**3)</td>
</tr>
<tr>
<td>O2lim</td>
<td>DO limitation term</td>
</tr>
<tr>
<td>OUR</td>
<td>oxygen uptake rate (M/L**3/T)</td>
</tr>
<tr>
<td>P</td>
<td>COD/VSS ratio</td>
</tr>
<tr>
<td>PH</td>
<td>pump head (Force/L**2)</td>
</tr>
<tr>
<td>Qi</td>
<td>flow rate into the wet well (L**3/T)</td>
</tr>
<tr>
<td>Qo</td>
<td>outlet flow from the wet well (L**3/T)</td>
</tr>
<tr>
<td>rdnb</td>
<td>nitrobacter decay rate (l/T)</td>
</tr>
<tr>
<td>rdns</td>
<td>nitrosomonas decay rate (l/T)</td>
</tr>
<tr>
<td>rnb</td>
<td>rate of nitrobacter production (M/L**3/T)</td>
</tr>
<tr>
<td>rnh4</td>
<td>rate of nitrosomonas production (M/L**3/T)</td>
</tr>
<tr>
<td>rns</td>
<td>rate of ammonia disappearance (M/L**3/T)</td>
</tr>
<tr>
<td>ro</td>
<td>autocorrelation</td>
</tr>
<tr>
<td>rsbs</td>
<td>rapidly biodegr. substr. utilization rate (M/L**3/T)</td>
</tr>
<tr>
<td>rsbp</td>
<td>slowly biodegr. substr. utilization rate (M/L**3/T)</td>
</tr>
<tr>
<td>rsd</td>
<td>rate of active mass production from Sd (M/L**3/T)</td>
</tr>
<tr>
<td>rsd'</td>
<td>rate of stored mass production from Sd (M/L**3/T)</td>
</tr>
<tr>
<td>Rsd</td>
<td>Direct growth rate coefficient (M/L**3/T)</td>
</tr>
<tr>
<td>Rt</td>
<td>kinetic parameter</td>
</tr>
<tr>
<td>rxa</td>
<td>active mass production rate (M/L**3/T)</td>
</tr>
<tr>
<td>Rxa</td>
<td>maximum specific growth rate (l/T)</td>
</tr>
<tr>
<td>rxd</td>
<td>organism decay rate (M/L**3/T)</td>
</tr>
<tr>
<td>rxe</td>
<td>nonbiodegradable residue production rate (M/L**3/T)</td>
</tr>
<tr>
<td>rxi</td>
<td>inert mass production rate (M/L**3/T)</td>
</tr>
<tr>
<td>Rxi</td>
<td>saturation coefficient</td>
</tr>
<tr>
<td>rwp</td>
<td>rate of stored mass utilization (M/L**3/T)</td>
</tr>
<tr>
<td>rxs</td>
<td>overall stored mass production rate (M/L**3/T)</td>
</tr>
<tr>
<td>S</td>
<td>substrate concentration (M/L**3)</td>
</tr>
<tr>
<td>Sbp</td>
<td>slowly biodegr. substrate concentration (M/L**3)</td>
</tr>
<tr>
<td>Sbs</td>
<td>rapidly biodegr. substrate concentration (M/L**3)</td>
</tr>
<tr>
<td>SCOUR</td>
<td>specific oxygen uptake rate (M/L**3/T)</td>
</tr>
<tr>
<td>Sd</td>
<td>soluble biodegr. substrate concentration (M/L**3)</td>
</tr>
<tr>
<td>SH</td>
<td>system head (Force/L**2)</td>
</tr>
<tr>
<td>Snh4</td>
<td>ammonia limitation coefficient</td>
</tr>
<tr>
<td>Sno2</td>
<td>nitrite limitation coefficient</td>
</tr>
<tr>
<td>SPEED</td>
<td>speed of variable speed pump (rpm)</td>
</tr>
<tr>
<td>SWD</td>
<td>sidewater depth (L)</td>
</tr>
<tr>
<td>sz**2</td>
<td>variance</td>
</tr>
<tr>
<td>t</td>
<td>time</td>
</tr>
<tr>
<td>TSS</td>
<td>total suspended solids (M/L**3)</td>
</tr>
<tr>
<td>v</td>
<td>velocity (L/T)</td>
</tr>
<tr>
<td>Vb</td>
<td>bulk fluid velocity (L/T)</td>
</tr>
<tr>
<td>V0</td>
<td>Stokes settling velocity (L/T)</td>
</tr>
<tr>
<td>VSS</td>
<td>volatile suspended solids concentration (M/L**3)</td>
</tr>
<tr>
<td>Vu</td>
<td>underflow velocity (L/T)</td>
</tr>
<tr>
<td>Xa</td>
<td>active mass concentration (M/L**3)</td>
</tr>
</tbody>
</table>
List of Symbols (cont'd)

\[ X_e \] - effluent solids concentration (M/L**3)
\[ X_i \] - inert mass concentration (M/L**3)
\[ X_{nb} \] - Nitrobacter concentration (M/L**3)
\[ X_{ns} \] - Nitrosomonas concentration (M/L**3)
\[ X_p \] - biodegr. particulate mass concentration (M/L**3)
\[ X_s \] - stored mass concentration (M/L**3)
\[ X_t \] - total mass concentration (M/L**3)
\[ Y_1 \] - yield coefficient for active mass (M/M)
\[ Y_2 \] - Yield coefficient for inert mass (M/M)
\[ Y_h \] - Yield coefficient in terms of COD
\[ z \] - Distance (L)
\[ z(t) \] - Observation at time t
1. INTRODUCTION

1.1 System Description

The purpose of this study is to examine the activated sludge process from a systems engineering perspective. Systems engineering has been defined by Luyben (1) as being the concept of examining the many parts of a complex plant together as a unit, with all the interactions included. All the plant units which affect the activated sludge process are shown in Figure 1.1. These units are:

- Influent pump station
- Air supply and distribution system
- Biological reactor
- Solids-liquid separator

The scope of this work is limited and defined by the system boundaries as shown in Figure 1.1. Everything outside the system boundaries is defined as "the outside world" and is connected to the system through the inputs and outputs.

The inputs to the system are:

- Influent wastewater from the sewer network. The composition and concentration of the wastewater, and the magnitude and variation of the flow, are dependent upon the sewer network and obviously affect the
Figure 1.1 Block diagram of the system
operation of the process. For example, the temperature of the wastewater affects the rate of the biological reactions, especially nitrification, and also determines the saturation value for the dissolved oxygen concentration. Changes in wastewater temperature may therefore have a significant effect on process performance.

- Environmental variables. Wind and air temperature affect the hydraulic regime in the solids-liquid separator and may affect clarification. Power consumption by the compressors is directly proportional to the mass flow rate of air, and is therefore affected by the humidity and temperature of the ambient air.

The significant inputs must be taken into account in the design of any control strategy. Many inputs are uncontrollable. However, they can often be measured and, in the case of the influent wastewater flow rate, predicted ahead of time with some degree of certainty.

The outputs of the system represent a link between the system to be studied and other systems - the solids processing train, chlorination, and the receiving body of water. System outputs will only be examined as they exert constraints on the system examined herein.
1.2 Statement of the Problem

Activated sludge wastewater treatment plants can be described as large complex systems with inputs which are not always predictable, and with strong interactions among process units. Some variables crucial to plant operation, such as characteristics of the influent wastewater or the activity of the microorganisms within the system, are difficult to measure on line. There are marked variations in influent wastewater flow rate, composition, and concentration. The treatment plant must be able to accommodate these changes. In the past, this problem has been primarily handled by increasing the capacity of the plant. An attractive alternative is the implementation of automatic process control. This approach is widely used in other continuous processes, such as petroleum refining.

Traditionally, activated sludge plants are designed using methods based primarily on steady state criteria. In practice, the process is dynamic. The problem, therefore, is not only to design control strategies that would improve operation of the process, including reductions in operating cost, but also to establish design requirements from the control engineer's point of view.

A logical question would be whether it is possible to improve operation or reduce operating costs by the use of process control. There are several characteristics of the activated sludge process that suggest that these objectives
may be accomplished by control:

- large throughput

- large variations in input (highly dynamic)

- complex mechanisms are involved

Processes with the above mentioned characteristics may also be difficult to control manually.

There are several problems, from a control engineer's point of view, that make the activated sludge process difficult to control. Olsson (2) has listed several such problems as:

- large variations in influent flow rate, composition, and concentration

- transport lags which can cause stability problems and degrade the performance of controllers

- reliable sensors are not available for several of the crucial variables. Included among these are many influent wastewater characteristics and the "activity" of the activated sludge.

- the time constants of the system vary greatly and range from minutes for the air supply/dissolved oxygen system to days for endogenous respiration. This can cause problems with numerical instability in computer simulations but may also be useful for uncoupling control loops when the time constants are substantially
different.

- most of the equations for describing the process are non-linear

- there are strong interactions among the process units

The applications of process control in wastewater treatment has lagged behind its application in other fields. Not all of the reasons for this are of a technical nature. Wastewater treatment does not have the same "driving force" as industry. Industry is driven by efforts to maximize profits and their survival is often based on their ability to do so. Wastewater treatment does not result in a profit and thus does not have the same "driving force" as industrial process operation. It is therefore important to identify other benefits of process control in wastewater treatment. Andrews (3) has listed the potential benefits of using dynamic models and process control in wastewater treatment as:

- Performance. Improved effluent quality can be achieved, as has been reported by West(4) and Joyce et al. (5), among others

- Productivity. An increase can be accomplished in the amount of waste that can be treated per unit of process capacity. Torpey (6) and Gould(7), among others, have reported improvements in productivity as a result of modified operational procedures.
- Reliability. Process control can significantly reduce the possibility of process failure.

- Stability. Mathematical models in combination with modern control theory can be valuable tools in examining and improving the stability of operation.

- Operating personnel. Dependable automatic control systems can be used to minimize the need for manual control.

- Operational cost. The cost of energy, manpower, and chemicals can be reduced.

- Start-up procedures. Start-up is obviously dynamic in nature and a knowledge of process dynamics could thus decrease the time required for start-up both when the plant is initially brought on-line or during recovery from a process failure.

- Development of manual operation guides. Knowledge of plant dynamics would be useful for this purpose in plants which do not have provisions for automatic control.

- Dynamic operation. It is frequently assumed that steady state operation is best for maximizing performance. However, this is not always so and dynamic models could be used to explore the possibility
of operating a process in a dynamic fashion so as to maximize performance.

- Variable efficiency operation. A treatment plant could be operated at variable efficiency in order to match the assimilative capacity of the receiving body of water which usually varies with time. This would require a good knowledge of the dynamic behavior of both the treatment plant and receiving waters and could be accomplished through the use of modern control systems.

The major thrust of this work is oriented toward the development of integrated control systems for the activated sludge process. The problem is defined from a control engineer's point of view, but with the specifics of the system kept always in mind. The solution to the problem can be broken into the following tasks:

- Objective function definition. Control of the process is achieved by minimizing the objective function.


- Controller design. This task includes selection and implementation of the appropriate control algorithms and strategies.
Each of these tasks will be examined first for each process, and then for the system shown in Figure 1.1.
1.3 General Approach to the Problem

A comprehensive dynamic model was developed for the system illustrated in Fig. 1.1. This model was used to represent the system with models for the control systems being superimposed on the plant model. Control strategies and interactions were examined using computer simulation. Control strategies were tested and compared with each other, and with a no-control case. Models for each process were in general selected from the literature although in some cases modifications were necessary for the purpose of this study. Model development is presented in Chapter 2.

It is important to note that in the simulations the controllers did not use any information that would not be available from the sensors in an actual plant. Although information about the concentration of active organism mass, substrate concentrations at several points in the biological reactor, and organism growth rate was in the model, it was not used by any of the proposed controllers. Only the variables that could be measured in a real situation, such as the dissolved oxygen concentration, were "known" by the controllers. Predictions of the influent flow rate were done beyond the sample, meaning that they were true predictions, and not just data fitted to autoregressive or moving average equations.

A limited amount of data was available from a real plant. This plant was the Sagemont plant in Houston, Texas. The
data used included effluent flow rates and effluent suspended solids concentrations. The pump station and the air supply and distribution system were also modeled after the Sagemont plant. However, the Sagemont plant was lightly loaded and thus did not exhibit strong dynamic behavior which prevented these data from being used more extensively. A short description of the Sagemont plant is presented in Appendix A. Complete information about this plant and the experimental work performed can be found in reports by Andrews (8-17).
1.4 Objectives

The primary objective of this research was to examine and define control strategies for the activated sludge process with consideration of the interactions between process units. The traditional approach to control (18,19,20,21) did not seem to be the best for this process. The research reported by Beck (22,23), Tong (24), Joyce et al. (5), Olsson (2,25,26,27), and Ortman et al. (28) provided the basis for the approach taken here. The task of defining and exploring control strategies includes:

- Research on state estimation methods
- Establishing process performance indices
- Controller design

Other objectives of this investigation were:

- to combine the use of deterministic and stochastic models in the development of an overall model for the process
- expand and adapt existing dynamic models for process units so they can be used in process control evaluation
- investigate process and control loop interactions
- examine interactions between plant design and operation
2. MODEL DEVELOPMENT

The control strategies are tested and evaluated using a mathematical model of the process. Development of new mathematical models for the process units was not, however, an objective of this study. Existing models were adopted from the literature whenever possible. However, in some cases it was necessary to modify the models so that they would be better suited for the purpose of this research.

2.1 Approach to Modeling

This chapter will not present a review of all the models that have been proposed for the activated sludge process but will instead describe the approach taken in some of the state of the art models for each process unit. Following this, the models will be adapted and combined for process control studies.

Considerable research has been devoted to modeling the activated sludge process. There are both stochastic (29,30) and deterministic (31,32,33) models in existence many of which have only been tested by computer simulations and comparisons with observations from the literature. These models may serve multiple purposes:

- Provide a better understanding of the mechanisms involved
- Serve as a quantitative technique for formulating hypotheses and guiding experimental research

- Permit exploration of possible control strategies for the process

- Suggest needed changes in design and/or operation for improving process performance and reliability or reducing operating costs.

In trying to attain the first two goals, the models tend to become more and more complex. Attempts have been made to use mechanistic models reduced in order and complexity for control applications (34, 35); however, much additional research is needed in this area.

There is a tendency in mechanistic modelling to underestimate effects of uncertainties in the model and the measurements. Another problem is estimation of the large number of parameters that are quite often present in such models. On the other hand, purely stochastic models do not utilize a priori knowledge about the system, and thus do not take advantage of "certainties" in the process. Both mechanistic and stochastic models will be considered in this study, and an effort will be made to blend these into a combined system model.
2.2 Influent Flow Prediction

Influent flow rate is one of the most important measurements in wastewater treatment and is measured in almost all plants regardless of size. It is a simple, reliable measurement and can be used to approximate the organic load to the process. Other variables that constitute the load to the process, such as the concentration and composition of organics in the wastewater, cannot be measured as easily.

Variation in the influent flow, which can be significant, depends on the sewer network connected to the plant. Substrate concentration variation is usually in phase with the changes in influent flow, thereby accentuating the variation in plant loading. Variations are larger in small to midsize plants. At a small plant such as the Sagemont plant, the flow rate ranged between 80 and 450 cu. m/hour on a typical dry weather day, whereas at Houston's larger 69 Street plant it varies between 9,000 and 16,000 cu. m/hour. The effects of such dynamic inputs can be observed in process variables such as dissolved oxygen concentrations (DO), mixed liquor suspended solids concentrations (MLSS), etc.
2.2.1 Literature Review

Prior knowledge of the influent flow pattern would be useful for control purposes in that appropriate control action could be taken in advance of detrimental effects. This has prompted researchers to explore the forecasting methods.

Stenstrom (33) used a finite Fourier series and field data to simulate an influent flow pattern. He superimposed a random noise component on this pattern in order to test plant control strategies. The objective of one of the control strategies proposed by Stenstrom was to maintain a constant growth rate in the activated sludge process. In this strategy specific oxygen uptake rate (SCOUR) was used as a dynamic indicator of the growth rate. Computer simulations indicated that the use of predicted values for the influent flow rate decreased the variability of SCOUR by 48% as compared to a control strategy without the flow prediction.

Shioja (38) applied multivariate analysis to the influent flow and obtained predictions one hour ahead as part of a control strategy for a pump station. Shioja used rainfall data in combination with the flow signal. Predictions were found to be inaccurate if the rainfall data were not used.

Much of the research on the application of stochastic modeling in wastewater treatment uses ARIMA time series models based on the methods of Box and Jenkins (39). In the
introduction to their text, they identify three areas in which time series models are particularly applicable:

- Obtaining predictions. This consists of forecasting of future values of a time series. Future values are obtained using both past and present values.
- Determining the transfer function of the system
- The design of control systems

One of the first efforts to obtain forecasts of influent flow rates was that of Goel and La Grega (36). They used a seasonal ARIMA model to predict, 24 hours in advance, average hourly flows to a conventional activated sludge process. A similar approach has been taken by Beck (37) in his study concerning identification and adaptive prediction of urban sewer flows.

In this research, the Box-Jenkins method was used only for forecasting of influent flows. Furthermore, only univariate models were considered. There are examples of the application of ARIMA models to determining transfer functions, primarily in the studies by Berthouex and coworkers (30,40). A recent study by Reagan (41) contains an excellent overview of applications of this method to wastewater treatment. The study points out a common problem in time series analysis applications, also observed by the author, that in many cases "predictions" are actually fits of a model to the data. Quite often the same data are used for estimating parameters for the model and for testing the
accuracy of the model. Therefore, the "predictions" are made within the same data sample and thus do not test the model as adequately as predictions made beyond the range of the data collected.

A simple technique for obtaining predictions of influent flow rates has been proposed by Dold (42). Dold used average flows for each time of day and stored these in a small computer. Average flows were updated by new measurements using the following expression:

\[ X_a'(t) = A \times X_m(t) + (1.0 - A) \times X_a(t) \]  \hspace{1cm} (2.1)

\[ X_a'(t) = \text{new average flow at time } t \]
\[ X_m(t) = \text{measured flow at time } t \]
\[ X_a(t) = \text{previous average at time } t \]
\[ A = \text{constant} \]

This method is commonly used in chemical engineering practice. Dold used these predictions to control an equalization tank and thus minimize variation of flow rate to a treatment plant. Although not as sophisticated as the Box-Jenkins approach, Dold has applied this method at full scale with resulting improvements in plant operation.
2.2.2 Proposed Model

In order to develop a mathematical model for the prediction of influent flow rate, data were obtained from the Sagemont wastewater treatment plant. Only data for the effluent flow rate were available, so it was assumed that this was approximately equal to the influent flow rate. However, this two flows will differ somewhat because of the dampening effects of the plant.

The effluent flow rate was measured continuously and recorded in the computer data base every minute. The average hourly flow rate was retrieved from these data for a period in the spring of 1983. When daily plots of effluent flow rate were visually examined, a consistent difference in daily flow pattern, as would be expected, was observed between weekdays and weekends. A decision was made to treat these separately. The flow pattern on weekends, in general, showed a more steady flow rate with less variation during the day and slightly higher averages.

The same methods can be used for flow prediction on weekends and weekdays, but the data base containing the history of flow on weekends is used in the first case, and that for flow on weekdays in the second case. Only the results obtained from the time series containing weekdays will be discussed here since the methods used for both predictions are identical. Since daily flow patterns showed more variation during weekdays than weekends, better results can
be expected for weekend data.

The time series used for the model development consisted of 55 weekdays in the spring of 1983, each day having 24 values of hourly average flow. These data were analyzed with computer programs that used the International Mathematical and Statistical Library (43) subroutines for time series analysis. These subroutines are based on the techniques developed by Box and Jenkins.

Figure 2.1 shows the autocorrelation and partial autocovariance plots obtained for the time series. The covariance between observation $z$ at time $t$, $z(t)$ and its value $z(t+k)$, separated by $k$ intervals of time, is called the autocovariance at lag $k$ and is defined by:

$$\gamma(k) = \text{cov}[z(t), z(t+k)] = E[(z(t) - \mu)(z(t+k) - \mu)]$$

$$\gamma(k) = \text{autocovariance at lag } k$$

$$\mu = \text{mean of the stochastic process}$$

$$E = \text{expectation}$$

$$\text{cov} = \text{covariance}$$

The autocorrelation at lag $k$ is defined as:

$$\rho(k) = \frac{E[(z(t) - \mu)(z(t+k) - \mu)]}{sz^{**2}}$$

$$\rho(k) = \text{autocorrelation at lag } k$$

$$sz^{**2} = \text{variance of the signal}$$
Figure 2.1 Time series analysis of flow rate data

Nondifferenced time series
Examination of these plots reveals that both the autocorrelation function and the partial autocovariance function have significant contributions at higher order lags. The practical implications of these contributions are that low order autoregressive models or moving average models would not be adequate to represent the time series. In order for the autoregressive model to be applicable, the partial autocovariance function should die out after a few lags. The autocorrelation function should not have significant contributions at higher lags for a moving average model to be applicable. The autocorrelation plot shows significant frequencies at lags of 12 and 24 hours, which was expected considering the periodicities normally present in domestic wastewater flow rates.

Since the data appeared non-stationary, an attempt was made to improve this by using a time series differenced by a lag of one hour. This approach was not effective. No improvement was achieved; on the contrary there was even more noise and no significant peaks. Plots for the autocorrelation and partial autocovariance for the time series differenced by one hour are presented in Figure 2.2.

Since peaks were observed at lags of 24 hours (Figure 2.1), and since the flow pattern has a 24 hour period, another attempt was made using a time series differenced by a lag of 24 hours. Results were still not much improved from the nondifferenced time series (Figure 2.3). It was observed
Figure 2.2 Time series analysis of flow rate data

Time series differenced by lag of 1 hour
Figure 2.3 Time series analysis of flow rate data
Time series differenced by lag of 24 hours
that peaks appeared again at lags of 24 hours, this time unexpectedly.

The next attempt was to use the time series differenced at lags of 12 and 24 hours, with results shown in Figure 2.4. Apparently too much filtering has been imposed on the data, and frequencies not present in the original data were forced onto the signal. Slutski (44) has shown that such a situation may arise when smoothing is repeatedly applied to a sequence of random variables.

An effort was then made to extract the maximum amount of deterministic information from the signal. The data from the ten most recent weekdays were used to calculate the average flow rate for each hour of the day. A new time series was then formed containing the perturbations from these averages, i.e., the differences between the average flow for a particular hour of the day and the actual, measured flow. This procedure was performed for each hour, with the most recent 240 hours being used to obtain average flows and the perturbations from the average pattern. The plots of the autocorrelation and the partial autocovariance functions are presented in Figure 2.5. It can be seen from the autocovariance plot that significant peaks occur at lags of 1 and 2 hours, with contributions at higher lags in the noise range. It should, therefore, be possible to achieve accurate predictions using an autoregressive model with lags of 1 and 2 hours.
Figure 2.4 Time series analysis of flow rate data

Time series differenced by lags of 12 and 24 hours
Figure 2.5 Time series analysis of flow rate data

Time series consists of perturbations from the mean
Several autoregressive models were tried and the results of flow predictions are presented in Figures 2.6, 2.7, and 2.8. Figure 2.6 shows the results of the prediction where the following expression was used to forecast the flow:

\[ F(t+1) = A + B*F(t) + C*F(t-1) + D*F(t-24) + E*F(t-48) \]  \hspace{1cm} (2.4)

Where:
- \( F(t) \) = flow rate at time \( t \)
- \( F(t-n) \) = flow rate at \( n \) sampling intervals before \( t \)
- \( A, B, C, D, E \) = parameters

The data base consisted of 55 weekdays of recorded flow rate. Since the most recent 240 hours of flow data were used to calculate parameters \( A, B, C, D \) and \( E \), predictions were made for 45 days. The data base for calculation of the parameters was continuously updated, and therefore the predictions were made beyond the sample and did not simply represent fits to equations. The best and the worst predictions are shown in Figure 2.6. The worst prediction occurred during a storm day when flow rates were unusually high.

Figure 2.7 shows the results of predictions made in a similar manner to those shown in Figure 2.6, except that lags at 1, 2 and 24 hours were used instead of 1, 24 and 48. Autoregressive-Moving Average models (ARMA) were also tried, but did not improve the predictions.
Figure 2.6 Flow prediction from nondifferenced time series
One step ahead, beyond the sample prediction
Figure 2.7 Flow prediction from nondifferenced time series

One step ahead, beyond the sample prediction
Figure 2.8 Flow prediction using perturbations

One step ahead, beyond the sample prediction
Figure 2.8 shows the most successful predictions of flow rates. Perturbations from the calculated mean flow pattern were used as a time series and the predictions were simply added to the mean flow value for the hour. It was observed that the predictions were quite accurate most of the time. The average error for predictions shown in Figure 2.8 (dry weather day) was 28 cu.m/hour which represents 10% of the average daily flow rate. During storm conditions predictions were less accurate. Predictions might be further improved by excluding storm days from the ten day historical data base as has been suggested by Olsson (45), or by including rainfall data and making predictions.
2.3 Pump Station

A mathematical model of the pump station was developed using data obtained from the pump station at the Sagemont wastewater treatment plant. The station was equipped with three pumps, each of which could be controlled in a different manner. A cross-sectional view of the pump station is presented in Figure 2.9 with the location of some of the instrumentation indicated. A detailed description of the pump station can be found in the report by Andrews (14). The control systems, all of which were based on liquid level measurement in the wet well, are:

- Variable speed pumping
- Automatic throttling of the pump discharge
- On/Off fixed speed pumping

All three pumps were identical except that the impellers had been trimmed on both the variable speed pump and the automatic throttling valve pump. Pump curves obtained from the manufacturer (46) were used in the model and are shown in Figures 2.10 and 2.11.

A mathematical model was used to generate equations to approximate the pump characteristic curves (Figures 2.10 and 2.11), and the system head curve. The following expression was used to describe the system head curve:

\[ SH = A + B \times \text{Flow}^{2} \]  

(2.5)
Figure 2.9 Cross-Sectional view of the Sagemont pump station
Figure 2.10 Pump performance curve - impeller not trimmed
Worthington 16MNZ
Centrifugal Pump
* Speed-875 RPM
Impeller Dia.-16"

Figure 2.11 Pump performance curve – trimmed impeller
where:

\[ SH = \text{system head, ft} \]

\[ A, B = \text{parameters} \]

\[ \text{Flow} = \text{Flow through the pump, gpm} \]

The fluid velocity through the pipes is calculated for a specific flow rate. Friction losses are computed and added into an overall coefficient for piping between the wet well and bar screen area. Friction losses are added to static head, and the procedure is repeated for a range of flows, generating data for the system head curve for a given valve opening. The IMSL (43) subroutine for least squares data fitting was used to determine parameters A and B in equation 2.5. Curves generated for valve openings of 25, 50, 75, and 100% open are presented in Figure 2.12. The listing of the program used to generate data for these curves is included in Appendix B. It can be observed from Figure 2.12 that the valve was oversized in that very little flow rate control could be accomplished by throttling the valve between 50 and 100% open which is the range of flows over which the pump operated. It can also be seen that the static head was quite high compared to the dynamic head.

The pump curves presented in Figure 2.10 and Figure 2.11 were approximated by the following expression:

\[ PH = A_1 + A_2 \times \text{Flow} + A_3 \times \text{Flow}^{**2} + A_4 \times \text{Flow}^{**3} \quad (2.6) \]
where:

\[ PH = \text{pump head, ft} \]

\[ A1, A2, A3, A4 = \text{parameters} \]

\[ \text{Flow} = \text{flow through the pump, gpm} \]

Parameters \( A1, A2, A3, \) and \( A4 \) are linear functions of the impeller speed:

\[ A1 = -24.375 + 0.085 \times \text{Speed} \quad (2.7) \]

\[ A2 = 0.01467 - 0.003946 \times \text{Speed} \quad (2.8) \]

\[ A3 = 0.27333 + 0.0000066442 \times \text{Speed} \quad (2.9) \]

\[ A4 = -0.069399 + 0.00004325 \times \text{Speed} \quad (2.10) \]

where:

\[ \text{Speed} = \text{impeller speed in RPM} \]

Parameters were obtained from historical data for variable speed pump operation. IMSL subroutines for least squares data fitting were used for estimation of parameters in equations 2.7 through 2.10.
2.4 Biological Reactor

The model of the biological reactor used in this research is a modified version of existing models. It uses features of models previously proposed by Stenstrom (33), Clifft (31), and Dold et al. (32). All of these models are mechanistic and deterministic. The biological reactor is modeled as four completely stirred tanks in series. The reactor flow terms were included in the model for the simulations but have been omitted for simplification in the description of the model presented herein.

2.4.1 Carbonaceous BOD Removal

2.4.1.1 Literature Review

The first model considered was that of Stenstrom (33). It is one of the dynamic models proposed by Andrews and a series of co-workers (Bryant (47), Busby(48), Stenstrom (33), Clifft (31)). The block diagram for carbonaceous BOD removal proposed by Stenstrom (33) is shown in Figure 2.13. In this model, the biomass was structured into three components - active, stored, and inert mass. Stored mass consists of biodegradable colloidal and suspended material entrapped within the floc, and poly-beta-hydroxybutyrate or glycogen-like compounds stored internally in the cell. The rate of production of stored mass was described with the following expression:

\[ rxs = Rt \left[ fsh \left( \frac{S}{KS+S} \right) - fs \right] Xt \]  \hspace{1cm} (2.11)
Figure 2.13 Schematic of Stenstrom's model

Substrate

\[ \text{rxs} = \text{Rt} \left[ \text{fsh} \left( \frac{S}{Ks+S} \right) - \text{fs} \right] \text{Xt} \]

Stored Mass

\[ \text{rxa} = \text{Rxa} \left( \frac{\text{fs}}{(Kfs+fs)} \right) \text{Xa} \]

Active Mass

\[ \text{rxi} = Y2 \text{ Rxi Xa} \]

Inert Mass
where

\( r_{xs} \) = rate of stored mass production (mg/L hour)
\( f_{sh} \) = maximum fraction of storage products (\( X_a/X_t \))
\( f_s \) = fraction of stored mass (\( X_a/X_t \))
\( S \) = substrate concentration (mg/L)
\( K_S \) = saturation coefficient (mg/L)
\( X_s \) = stored mass concentration (mg/L)
\( X_a \) = active mass concentration (mg/L)
\( X_t \) = total organisms conc. \( X_a + X_s + X_i \) (mg/L)
\( X_i \) = inert mass (mg/L)

Active mass is formed from stored mass through synthesis. The following expression was proposed by Stenstrom for this reaction:

\[
 r_{xa} = R_{xa} \left( \frac{f_s}{(K_{fs} + f_s)} \right) X_a
\]  \hspace{1cm} (2.12)

where

\( r_{xa} \) = rate of active mass production (mg/L hour)
\( R_{xa} \) = maximum specific growth rate (1/hour)
\( K_{fs} \) = saturation coefficient

A first order reaction rate was assumed for the production of inert mass from active mass:

\[
 r_{xi} = Y_2 R_{xi} X_a
\]  \hspace{1cm} (2.13)

where

\( r_{xi} \) = rate of inert mass production (mg/L hour)
\( Y_2 \) = yield coefficient (\( mgX_i/mgX_a \))
\( R_{xi} \) = specific organism decay rate (1/hour)
Block diagrams showing carbonaceous BOD removal according to the models of Clifft (31) and Dold, et al. (32) are shown in Figures 2.14 and 2.15, respectively. Both of these models represent an improvement over that of Stenstrom's in that the influent biodegradable substrate is structured into two portions on the basis of the respective removal mechanisms. However, the models differ in their means of classification of influent substrate.

Clifft proposed the division of influent biodegradable substrate into soluble and particulate fractions. Substrate capable of passing through a glass fiber filter (Whatman No. 42) was arbitrarily defined as soluble substrate. Division of substrate into soluble and particulate fractions is an attempt to establish criteria for predicting substrate removal kinetics using a relatively simple physical measurement which is widely used in the field. It is assumed that the mechanisms for removal of soluble and particulate fractions are different.

Clifft assumes that soluble substrate is removed by two mechanisms occurring simultaneously, as noted in Figure 2.14. These mechanisms are as follows:

a) Direct synthesis of active mass

\[ rsd = Rsd * Xa * Sd \]  \hspace{1cm} (2.14)

\[ rsd = \text{rate of production of active mass} \]
Figure 2.14 Structure of Cliff's model

\[ \frac{dS_D}{dt} = -k_S S_D (1 - \frac{C}{S}) \]
\[ \frac{dx_A}{dt} = \frac{k}{1 + (x/S)} \]
\[ \frac{dx_I}{dt} = -k_D x_A \]
\[ \frac{dx_S}{dt} = k_P x_A \]
\[ \frac{dx_P}{dt} = -k_P x_A \]

Where:
- \( S_D \) is the soluble substrate
- \( x_A \) is the active mass
- \( x_I \) is the inert mass
- \( x_S \) is the stored mass
- \( x_P \) is the particulate substrate
- \( k \) is the rate constant
- \( k_S \) is the solubility constant
- \( k_P \) is the particulate uptake rate
- \( C \) is the concentration of the limiting substrate
from $S_d$ (mg/L hour)

$S_d = \text{Biodegradable soluble substrate concentration (mg/L)}$

$R_{sd} = \text{Direct growth rate coefficient (L/mg hour)}$

b) Formation of stored mass

$$r_{sd}' = -K_t \times X_a \times S_d \times [f_{shat} - f_s] \quad (2.15)$$

$r_{sd}' = \text{rate of production of stored mass from } S_d$ (mg/L hour)

$K_t = \text{Transport Rate coefficient (L/mg hour)}$

$f_{shat} = \text{Maximum fraction of stored mass}$

$f_s = \text{fraction of stored mass, } X_s/(X_s + X_a)$

Particulate substrate is assumed to become trapped in the floc instantaneously. Once it is a part of the floc, it is gradually hydrolyzed by the organisms. The proposed rate equation for hydrolysis is:

$$r_{xp} = -R_h \times [f_p/(K_{sp} + f_p)] \times X_a \quad (2.16)$$

where

$r_{xp} = \text{rate of stored mass utilization (mg/L hour)}$

$X_p = \text{biodegradable particulate substrate concentration (mg/L)}$

$R_h = \text{Hydrolysis rate coefficient (1/T)}$

$f_p = \text{Fraction of stored particulate substrate } X_p/(X_p + X_a)$

$K_{sp} = \text{Saturation coefficient (dimensionless)}$
For low (less than 5 days) sludge ages, Clift used a simplified version of equation (2.8):

\[ r_{xp} = -K_p \times X_a \quad (2.17) \]

\(K_p\) = Hydrolysis rate coefficient (1/hour)

Stored mass is formed from both soluble and particulate substrates. The overall rate expression for stored mass is:

\[ r_{xst} = K_t \times X_a \times S_d \times [f_{shat} - f_s] + K_p \times X_a - R_{xa} \times X_a \times f_s / Y_l \quad (2.18) \]

\(R_{xa}\) = growth rate coefficient 1/hour

\(Y_l\) = mass of \(X_a\) produced per unit mass of \(X_s\) or \(S_d\) utilized

Active mass is formed through synthesis of soluble and/or stored mass. Active mass is then converted to inert mass by organism decay. These reactions lead to the rate equation for formation of active mass:

\[ r_{xat} = R_{xa} \times X_a \times f_s + Y_l \times R_{sd} \times X_a \times S_d - K_d \times X_a \quad (2.19) \]

All concentrations in both Clift's and Stenstrom's models are expressed in terms of oxygen equivalents.

The block diagram for the model proposed by Dold, Ekama, and Marais (32) is shown in Figure 2.15. In this model substrate is expressed in units of oxygen equivalents (COD) and biomass is in units of VSS. Influent biodegradable substrate is structured into two fractions based on the rate
\[ \frac{dS_{bs}}{dt} = \left[ \frac{K_{ms} S_{bs}}{K_{ss} + S_{bs}} \right] x_a \]

\[ \frac{dx_a}{dt} = \left[ \frac{K_{mp} (x_a / x_a)}{(K_{sp} + P) + (x_s / x_a)} \right] x_a \]

\[ \frac{dx_a}{dt} = -b_h x_a \]

\[ \frac{ds_{bp}}{dt} = P(1-f)b_h x_a \]

\[ \frac{dx_e}{dt} = f_b x_a \]

Figure 2.15 Structure of University of Cape Town model
at which it is utilized - rapidly biodegradable and slowly biodegradable substrate. Dold et al. conducted experiments where a daily cyclic square wave loading was introduced to a bench scale biological reactor. Figure 2.16 shows the response in oxygen uptake rate (OUR) when three different substrates were fed to the reactor. The substrates were glucose, soluble starch and domestic wastewater. Glucose was used as the rapidly biodegradable substrate and starch was used to represent a slowly biodegradable substrate.

The rapidly biodegradable substrate is utilized according to the following rate expression:

\[
rsbs = -\left[\frac{Kms \cdot Sbs}{(Kss + Sbs)}\right] \cdot Xa
\]  \hspace{1cm} (2.20)

\(Sbs\) = rapidly biodegradable substrate concentration, mg COD/l

\(Kms\) = maximum specific growth constant utilizing soluble substrate, mg COD/mg VSS .d

\(Kss\) = saturation coefficient for soluble substrate, mgCOD/l

\(Xa\) = active mass concentration, mg VSS/l

The rate of disappearance of the slowly biodegradable substrate is given by:

\[
rsbp = - Ka \cdot Sbp \cdot Xa \cdot (fma - Xs/Xa)
\]  \hspace{1cm} (2.21)

\(Sbp\) = slowly biodegradable substrate concentration, mg COD/L
a). Glucose

b). Soluble Starch

c). Domestic Wastewater

Figure 2.16 Cyclic OUR response to different substrate after Dold (2)
\( K_a = \) particulate substrate adsorption constant, 
\( L/mgVSS \text{ d} \)

\( f_m = \) maximum fraction of substrate (as VSS) that can 
be incorporated in the active volatile sludge 
mass, \( mgVSS/mgVSS \)

\( X_S = \) stored mass concentration, \( mg \text{ VSS/L} \)

Stored mass is formed directly from the slowly biodegradable 
substrate, and is synthesized into active mass:

\[
rx_s = \frac{(K_a S_b p X_a (f_m X_s / X_a))}{P} - \frac{[K_m p X_s (K_s p X_A + X_s P)]}{X_s} \quad (2.22)
\]

\( P = \) COD/VSS ratio \( mgCOD/mgVSS \)

\( K_m p = \) maximum specific substrate utilization 
rate constant for stored mass (as COD), 
\( mgCOD/mgVSS \text{ d} \)

\( K_s p = \) half saturation coefficient for the utilization 
of stored COD, \( mgCOD/mgVSS \)

A first order rate equation was proposed by Dold et al. for 
organism decay:

\[
rx_d = -b_h X_a \quad (2.23)
\]

\( b_h = \) endogenous respiration rate of heterotrophic 
organisms, \( 1/d \)

Organisms are assumed to form two fractions after lysis — 
nonbiodegradable residue \( X_e \), and slowly biodegradable
residue - Sbp. Both fractions are released into the liquid phase. The rate of production of nonbiodegradable residue is given by equation:

\[ r_{xe} = f \cdot b \cdot h \cdot Xa \quad (2.24) \]

- \( f \) = nonbiodegradable fraction of \( Xa \) mgVSS/mgVSS
- \( Xe \) = nonbiodegradable residue mgVSS/l

The slowly biodegradable fraction that is released into the liquid phase is:

\[ dSbp(*)/dt = P \cdot (1-f) \cdot bh \cdot Xa \quad (2.25) \]

and should be included in the overall equation (8) for slowly biodegradable substrate.

The overall active mass equation includes formation of active mass from both stored mass and rapidly biodegradable substrate and the decrease in active mass due to organism decay:

\[ r_{xat} = Yh \cdot [(K_{ms} \cdot Sbs)/(K_{ss} + Sbs) + K_{mp} \cdot Xsp/(K_{sp} \cdot Xa + Xsp)] \cdot Xa - bh \cdot Xa \quad (2.26) \]

- \( Yh \) = yield coeff. in terms of substrate as COD mgVSS/mgCOD

An on-line method for estimation of influent substrate structure using OUR is not currently available. However, the importance of OUR as a control parameter is recognizable from the results obtained with this model.
2.4.1.2 Proposed Model for Carbonaceous BOD Removal

The model for carbonaceous BOD removal used in this research is structured with respect to both influent substrate and sludge mass. All concentrations are expressed in oxygen equivalents. Both Clifft's and Dold's model distinguish between fractions of substrate that degrade at different rates. Clifft's division of substrate into soluble and particulate fractions was adopted here. The reason behind this choice is that filtration is a relatively easy and frequently used technique for estimating substrate structure. However, this approach implicitly assumes that the rate at which the substrate degrades is only related to the particle size. It should also be noted that filtering, as presently practiced, is not truly an on-line technique. If a method for estimating rapidly and slowly biodegradable fractions on line was available, Dold's model would be more applicable to control. Equation 2.14 was used for describing direct synthesis of active mass from soluble substrate.

Soluble substrate is incorporated in the floc phase (stored mass) according to equation 2.15, but the definition used by Stenstrom for the fraction of stored mass was used:

\[ fs = \frac{X_s}{(X_s + X_a + X_i)} \]  \hspace{1cm} (2.27)

Equation 2.17, describing utilization of particulate stored substrate, was modified as shown below:
\[ r_{xp} = -R_h \cdot X_p \cdot X_a \]  \hspace{1cm} (2.28)

The reaction is assumed to be first order with respect to both active mass and particulate substrate.

Active mass formation from stored mass was modeled by the expression:

\[ r_{xa} = R_{xa} \cdot (f_{sl}/(f_{sl}+1.0)) \cdot X_a \cdot O_{2lim} \]  \hspace{1cm} (2.29)

where

- \( f_{sl} \) = stored mass in active fraction \( X_s/X_a \)
- \( O_{2lim} \) = oxygen limitation term, \( DO/(K_{DO}+DO) \)

A first order expression (eq. 2.13) was used to describe organism decay.
2.4.2 Nitrification

The nitrification model developed by Poduska (49), also used by Stenstrom (33), was adopted. No modifications were made except for the oxygen limitation term imposed on growth of nitrosomonas and nitrobacter as proposed by Stenstrom and Poduska (50). The growth rate of nitrosomonas is described by:

\[ rns = mhns \times (\frac{nh4}{(snh4+nh4)}) \times O2lim \times Xns \]  
  \[ (2.30) \]

where:

- \( rns \) = rate of nitrosomonas production (mg/L hour)
- \( mhns \) = maximum growth rate coefficient for nitrosomonas (l/hour)
- \( nh4 \) = ammonia concentration (mg/L)
- \( snh4 \) = substrate limitation coefficient (mg/L)
- \( Xns \) = nitrosomonas concentration (mg/L)

The growth of nitrobacter is assumed to behave according to the following:

\[ rnb = mhnb \times (\frac{no2}{(sno2+no2)}) \times O2lim \times Xnb \]  
  \[ (2.31) \]

where:

- \( rnb \) = rate of nitrobacter production (mg/L hour)
- \( mhnb \) = maximum growth rate coefficient for nitrobacter, (l/hour)
- \( no2 \) = nitrite concentration (mg/L)
- \( sno2 \) = substrate limitation coefficient (mg/L)
Xnb = nitrobacter concentration (mg/L)

Both nitrosomonas and nitrobacter are assumed to undergo first order decay as shown in Equations 2.32 and 2.33.

\[ r_{dns} = k_{dns} \times Xns \] (2.32)
\[ r_{dnb} = k_{dnb} \times Xnb \] (2.33)

Ammonia is consumed during synthesis by heterotrophic organisms and nitrosomonas. Nitrogen is returned to solution as a result of organism decay. The final rate equation for disappearance of ammonia is:

\[ r_{nh4} = Kn \times rxaf - (1.0 - Y2) \times Kn \times rxi + rns/Yns \] (2.34)

where:

\( r_{nh4} \) = rate of ammonia disappearance (mg/L hour)
\( Kn \) = stoichiometric coefficient
\( rxaf \) = total rate of active mass formation (mg/L hour)
\( rxi \) = rate of inert mass production (mg/L hour)
\( rns \) = rate of nitrosomonas production (mg/L hour)
\( Y2 \) = yield coefficient

The total rate of active mass formation (\( rxaf \)) includes formation of active mass from stored mass (eq. 2.29) and direct synthesis of soluble substrate to active mass (eq. 2.14).
2.4.3 Dissolved Oxygen Dynamics

The dissolved oxygen concentration in a biological reactor is a function of the depletion caused by microbial activity and transfer from the gas phase. Dissolved oxygen depletion due to microbial activity can be described by the following equation:

\[
\text{OUR} = [(1-Y_1)/Y_1] \cdot r_{xa} + (1-Y_2) \cdot r_{xi} + 3.4 \cdot r_{ns}/Y_{ns} + 1.1 \cdot r_{nb}/Y_{nb}
\] (2.35)

where:

\text{OUR} = \text{oxygen uptake rate (mg/L hour)}

The transfer rate of oxygen from the gas to the fluid phase is given by the two film mass transfer equation of Lewis and Whitman (51).

\[
r_{do} = K_{la} \cdot (D_{os} - D_{o})
\] (2.36)

where:

\text{K}_{la} = \text{transfer rate coefficient (1/hour)}

\text{D}_{os} = \text{dissolved oxygen saturation concentration (mg/L)}

\text{D}_{o} = \text{dissolved oxygen concentration (mg/L)}

Procedures for estimating \text{K}_{la} have been examined by a number of researchers (52-56). In this study, the method used by Goto (57) for \text{K}_{la} estimation was adopted. This method uses a function relating air flow rate to \text{K}_{la}, and will be discussed in section 2.6.
2.5 Solids-Liquid Separator

The solids-liquid separator is an essential portion of the activated sludge process. There are three basic functions which the solids-liquid separator performs:

- Clarification of the effluent
- Thickening of the sludge
- Storage of sludge

There have been several mathematical models developed for the separator (58, 59, 47, 60, 61, 62). Some of the previous research on dynamic modeling of the separator will be first reviewed, and then the model proposed in this research will be presented.
2.5.1 Thickening Model - Literature Review

A dynamic model for the thickening function of the separator, as developed by Bryant (47) and subsequently modified by Stenstrom (33), will be presented in some detail since the model proposed herein was developed by an expansion of these models. The following is a description of Bryant's model.

The material balance for the separator can be described by the following expression:

\[ \frac{\partial C}{\partial t} = \frac{\partial (D\partial C)}{\partial Z^2} - \frac{\partial (VC)}{\partial Z} - \text{reaction} \quad (2.37) \]

where:

- \( C \) = concentration (M/L**3)
- \( t \) = time (T)
- \( D \) = dispersion (L**2/T)
- \( V \) = velocity (L/T)
- \( Z \) = distance (L)

The thickening model proposed by Stenstrom makes use of equation 2.37, often referred to as the continuity equation. Stenstrom's model was based on the following assumptions:

1. The dispersion is zero (Plug flow)

2. The mass flux into a differential volume cannot exceed the flux which the volume is capable of passing nor can it exceed the flux which the
next higher differential volume is capable of transmitting.

3. The settling velocity is a function only of solids concentration, except when #2 is violated.

4. The bottom of the separator represents a physical boundary to sedimentation; therefore, the settling flux at the bottom of the separator is zero.

5. The solids concentration is completely uniform in any horizontal plane.

6. The solids in the effluent may be predicted by an empirical relationship.

7. There is no biological reaction in the separator.

Stenstrom divided the settler into several finite elements, modeled each as a completely mixed tank, and wrote a mass balance around each tank. Using expression 2.37 and assumptions 1-7, he obtained the following expression for the change in concentration for the ith element:

\[ \frac{\partial c}{\partial t} = \nu u (C_i - C_i) / dz + (\text{Min}(G_{si}, G_{si-1}) - \text{Min}(G_i, G_{s+1})) / dz \quad (2.38) \]

where:

\[ \nu u = \text{bulk downward velocity (L/T)} \]
C = slurry concentration (M/L**3)
Gs = settling flux = Vs*C
i = subscript denoting ith element
dz = finite element height (L)
Min = minimum function

\[ f = \text{Min}(a,b) \quad f = a \text{ for } a < b \]
\[ f = b \text{ for } a > b \]

For determination of the interface settling velocity, Stenstrom refers to Coe and Clevenger (63) and Talmadge and Fitch (64). They determined the settling velocity by measuring the initial subsidence rate of a slurry in a one liter or larger cylinder. After repeating this procedure for several different concentrations, a settling velocity function is defined. This function was used to predict the interface settling velocity as a function of slurry concentration.

Hill (62) followed the same procedure using sludge taken from the return sludge line at the Sagemont plant. He fitted the data for settling velocity to the following expression developed by Vesilind (65):

\[ Vs = V_0 \exp(-b*C) \quad (2.39) \]

where:

Vs = settling velocity at concentration C (L/T)
C = concentration of sludge (M/L**3)
b = empirical parameters
\[ V_0 = \text{Stokes settling velocity for a single discrete particle (L/T)} \]

Parameters \( V_0 \) and \( b \) were obtained by plotting the settling velocity versus the natural logarithm of concentration and measuring the slope \( (b) \) and the intercept \( (V_0) \). Equation 2.39 and the experimental values for parameters \( V_0 \) and \( b \) obtained by Hill were used in this study.

Equation 2.38 is written for each finite volume except for the top and bottom elements. Equations for the top and bottom elements include the boundary condition and are given below:

**Top element:**

\[ \frac{dC_l}{dt} = \left( \text{FLUXIN} - U C_l - \min(Gs_1, Gs_2) \right) / dz \quad (2.40) \]

**Bottom element:**

\[ \frac{dC_n}{dt} = (U(C_{n-1} - C_n) + \min(Gs_n, Gs_{n-1}) / dz \quad (2.41) \]

where:

- \( \text{FLUXIN} = \text{net flux into the separator (M/T L**2)} \)
- \( n = \text{subscript of bottom element} \)

Hill (62) modified Stenstrom's model to take into account the conical shape of the lower part of the separator. The equation for the bottom element proposed by Hill thus differs from equation 2.41 in that the constraint is not imposed on the flux in the last element, therefore \( \min(Gs_n, Gs_{n-1}) \) is replaced by \( [Gs_n] \).
2.5.2 Clarification - Literature Review

Clarification has a strong impact on process performance, particularly since the effluent suspended solids (SS) usually represents a major portion of the effluent BOD. Several models have been proposed for clarification, all of which are empirical. Busby (48) used data obtained by Pflanz (66) and proposed the following model:

\[ X_e = K \times (F_0 + F_r) \times MLSS / Area \] (2.42)

where:

\( X_e \) = effluent suspended solids concentration (M/L**3)
\( K \) = constant
\( F_0 \) = influent flow to the plant (L**3/T)
\( F_r \) = recycle flow rate (L**3/T)
\( MLSS \) = concentration of solids entering separator (M/L**3).

\( Area \) = cross sectional area of the separator (L**2)

This model was also later used by Stenstrom (33). Another empirical model was developed in an EPA study (58) in which full scale experiments were conducted and the following model for effluent suspended solids was proposed:

\[ X_e = B_1 + B_2(F_0/A) - B_3*MLSS \] (2.43)

where:

\( B_1, B_2, B_3 \) = empirical parameters

Chapman (59), based on pilot scale experiments, suggested
that the effluent SS are affected by changes in influent flow rate to the solids-liquid separator and by changes in MLSS. He used a first order model to describe the response of effluent suspended solids concentration to a decrease in flow rate and/or to a change in MLSS. A second order model was used to describe the effect of increase in flow rate on effluent suspended solids. Chapman's equation is:

\[ X_e = K_1 + K_2 \times \text{MLSS} + K_3 \times F_{\text{ain}} + \text{SWD} \times (K_4 - K_5 \times F_{\text{ain}}) \]  

(2.44)

where:

- \( K_1 - K_5 \) = constants
- \( F_{\text{ain}} \) = Fins/Area
- Fins = flow rate into the separator (L**3/T)
- SWD = sidewater depth (L)

Equation 2.44 explained 78% of the variation in his data.
2.5.3 Proposed Model

2.5.3.1 Clarification

Two approaches are possible for modeling clarification phenomena. The first is physical, and is related to turbulence and hydraulic conditions in the separator. The second is biological and it is related to the settling properties of the solids. Both physical and biological mechanisms affect the process of clarification at the same time.

Research by Chapman (59) and Olsson and Chapman (67) examine in depth the physical aspects of the problem. They show that the effluent suspended solids are related to the hydraulics of the separator and that turbulence due to sudden changes in flow rate has detrimental effects. Their observations will be used in developing control strategies for improving the operation of the solids-liquid separator. These control strategies are based on minimizing changes in flow rate. However, the biological aspect of the problem, as they influence the settling characteristics of the sludge, also should be included in the model in order to fully describe clarification.

The biological aspects of the clarification problem are the result of the strong interaction between the biological reactor and the separator. Settling properties of the solids are influenced by conditions in the biological
reactor and therefore control of conditions in the biological reactor will affect the operation of the separator. This aspect of the problem will be discussed in Chapter 4.

A substantial amount of data collected at the Sagemont plant was available for this research. Effluent suspended solids were measured on-line for a period of several months and were recorded for later analysis. The suspended solids concentrations measured were unusually low being in the range of 2-4 mg/L. Several experiments were conducted during which hydraulic disturbances were introduced and changes in the effluent concentration examined. These experiments were performed and have been described in detail by Hill (62). Results from these experiments will be compared with routinely collected data over a period of several months.

Figure 2.17 shows forcings introduced during one of the experiments and the resulting response in effluent suspended solids. Signals are shown overlayed and seem to be correlated. The absolute change in the effluent solids is not large (5mg/L maximum).

One minute averages of effluent flow rate and suspended solids concentrations were recorded and plotted daily. Examination of records from several months of operation indicated that the correlation between the effluent flow rate and effluent suspended solids observed during short
Figure 2.17 Effluent flow rate and suspended solids data from Sagemont WTP, April 26, 1983.
term experiments did not exist on other days. Several examples will be presented in order to illustrate that this correlation was not very consistent.

Figure 2.18 contains data collected on April 24, 1983. The figure shows very good correlation between the signals, even at low flow rate (only the morning hours are shown, since the instruments were recalibrated in the afternoon). This would support the conclusion from the short term experiments that the effluent suspended solids concentration is strongly influenced by changes in flow rate. However, figures 2.19 and 2.20 from data obtained approximately a month earlier show little response in effluent suspended solids to changes in flow rate, even at higher magnitudes and amplitudes. There were a number of days both before and after the experiment that exhibited similar behavior.

Another contradiction is shown in Figures 2.21 and 2.22. The effluent flow rate signal has both a low mean and a low amplitude, but the effluent suspended solids signal shows wide variations and magnitudes up to two times greater than those created by hydraulic forcings during the experiment. One might speculate that it supports the statement made earlier in the text that biological, as well as physical, mechanisms affect clarification. There is also a strong possibility that the stochastic component of clarification overshadows the deterministic component at these low concentrations.
Figure 2.18 Effluent flow rate and suspended solids data from Sagemont WTP, April 24, 1983.
Figure 2.19 Effluent flow rate and suspended solids data from Sagemont WTP, March 26, 1983.
Figure 2.20 Effluent flow rate and suspended solids data from Sagemont WTP, March 20, 1983.
Figure 2.21 Effluent flow rate and suspended solids data from Sagemont WTP, May 29, 1983.
Figure 2.22 Effluent flow rate and suspended solids data from Sagemont WTP, June 1, 1983.
Several conclusions may be drawn from these data. The very low average values and amplitudes of the effluent solids concentrations which were observed most of the time indicate that the plant shows little dynamic behavior. The signal to noise ratio is also low when the signal is in the range of 2-5 mg/L indicating the presence of a strong random component.

As mentioned earlier in the text, the solids-liquid separator performs three functions - clarification, thickening and storage. The separator model used in this study does not include a model for clarification, therefore effluent suspended solids concentration is not predicted during normal operating conditions. However, the discharge of large quantities of effluent suspended solids in the event of a failure of the separator to perform thickening or storage functions during upset conditions is predicted by the thickening model. This model is used for developing and testing of control strategies. Although clarification is not modeled, it is qualitatively considered in the design of the control strategies.
2.5.3.2 Thickening

This model was developed from the model by Bryant (47), as modified by Stenstrom (33), and has been successfully used to predict underflow sludge concentrations. It can not, however, predict the sludge blanket level in the solids-liquid separator, and does not predict the solids concentration in the clear zone above the sludge blanket. In this research the model has been modified to include the region above the sludge blanket. Since it predicts the solids profile in the solids-liquid separator, the sludge blanket level can be predicted. A description of the model follows.

The solids-liquid separator is idealized as being divided into "m" layers as shown in Figure 2.23. Influent flow enters the separator at the feed point and is assumed to be instantly and completely mixed in layer "n". Three different regions are defined within the separator - the region above the feed point (layers 1 through n-1), the feed point (layer n), and the region below the feed point (layers n+1 through m).

Some of the terms that will be used in the model are defined as:

\[ V_b = \frac{F_{out}}{\text{Area}} \]  \hspace{1cm} (2.45)

where:

\( V_b \) = upward bulk fluid velocity (L/T)
Figure 2.23 Idealized Solids-liquid Separator
\[ F_{\text{out}} = \text{effluent flow rate from the separator (L}^3/T) \]
\[ \text{Area} = \text{cross-sectional area of the separator (L}^2) \]

and

\[ V_u = \frac{F_{\text{r}}}{\text{Area}} \quad (2.46) \]

where:
\[ V_u = \text{downward bulk fluid velocity (L/T)} \]
\[ F_{\text{r}} = \text{recycle flow rate (L}^3/T) \]

also:

\[ G_s(i) = V_s(i) \times c(i) \quad (2.47) \]

\[ G_s(i) = \text{solids flux due to concentration in layer } i \text{ (M/L}^2/T) \]
\[ V_s(i) = \text{settling velocity determined from settling test (L/T)} \]
\[ c(i) = \text{solids concentration in layer } i \text{ (M/L}^3) \]

and finally:

\[ G_{\text{up}}(i) = V_b \times c(i) \quad (2.48) \]

\[ G_{\text{up}}(i) = \text{upward movement of solids due to } V_b \text{ (M/L}^2/T) \]

It is assumed that the bulk fluid velocity equals \( V_b \) in layers 1 through \( n-1 \) and \( V_u \) in the layers \( n+1 \) through \( m \). Within layer \( n \) the bulk fluid is assumed to move up with a velocity \( V_n \).

\[ V_n = V_b - V_u \quad (2.49) \]

\[ V_n = \text{bulk velocity in layer } n \text{ (L/T)} \]
The profile of velocities assumed within the separator is shown in Figure 2.24. Studies by Crosby (68) showed that flow patterns in the separator are much more complex than assumed here, but this simplified velocity profile can be used to develop a model since the emphasis of this research is on control, rather than on model development. Several suggestions for making the model more realistic will be given later.

The region below the entrance (lower region) is modeled in a manner similar to that used by Stenstrom. In the region above the feed point it is assumed that the solids have to overcome the upward velocity, \( V_b \), in order to settle. Five groups of layers are defined in this model and equations describing changes in solids concentrations in each group of layers are given below.

Layer 1:

\[
z(1) \frac{dc(1)}{dt} = G_{up}(2) - G_{down}(1) - G_{up}(1) \tag{2.50}
\]

\( z(1) \) = height of layer 1

\( G_{down}(1) = G_s(1) \) if \( c(2) < C_t \)

\( G_{down}(1) = \min \{ G_s(1), G_s(2) \} \) if \( c(2) > C_t \)

Layers 2 - \( n-1 \):

\[
\begin{align*}
z(i) \frac{dc(i)}{dt} & = G_{up}(i+1) - G_{up}(i) \\
& + G_{down}(i-1) - G_{down}(i) \tag{2.51}
\end{align*}
\]
Figure 2.24  Idealized Bulk fluid velocity profile within the separator
\[ z(i) = \text{height of layer } i \ (L) \]
\[ i = 2 \text{ through } n-1 \]
\[ c(i) = \text{concentration in layer } i \]

\[ G_{\text{down}}(i) = G_s(i) \text{ if } c(i+1) < C_t \]

\[ G_{\text{down}}(i) = \min [G_s(i), G_s(i+1)] \text{ if } c(i+1) > C_t \]

It is assumed that \( C_t \) is a threshold where the settling flux in a given layer will start to affect the settling flux in the next upper layer. The top of the sludge blanket was defined as the closest layer to the top of the clarifier with solids concentration equal to or greater than \( C_t \).

Layer \( n \) (feed point):

\[ z(n)^* \frac{dc(n)}{dt} = (F_r+F_{\text{out}})^*MLSS + G_{\text{down}(n-1)} - G_{u}(n) - V_u*c(n) - \min[G_s(n), G_s(n+1)] \] \hspace{1cm} (2.52)

Layers \( n \) through \( m-1 \):

\[ z(i)^* \frac{dc(i)}{dt} = \min[G_s(i-1), G_s(i)] - \min[G_s(i), G_s(i+1)] - V_u*c(i) + V_u*c(i-1) \] \hspace{1cm} (2.53)

\[ i = n \text{ through } m-1 \]

The last (bottom) layer (\( m \)):

\[ z(m)^* \frac{dc(m)}{dt} = \min[G_s(m-1), G_s(m)] - V_u*c(m) + V_u*c(m-1) \] \hspace{1cm} (2.54)

According to solids flux theory, the ideal solids
concentration profile in the separator consists of three different concentration zones. Figure 2.25, taken from Stenstrom (33), presents the ideal and the expected (non-ideal) solids profile as predicted by solids flux theory. In order to make a comparison between model predictions and theory, a sinusoidal input for flow rate and concentration to the separator was simulated. Solids profiles obtained from simulations are shown for low and high loadings in Figure 2.26. At both loadings the solids profile in the separator is in qualitative agreement with the theoretical profile shown in Figure 2.25.

Several possible improvements to the model would be:

- Modify the assumption that the radial concentration profile is uniform through the layer thus allowing for a concentration gradient in two dimensions rather than one.

- Include the effects of turbulence on the transfer of solids between layers

- Structure the solids with respect to settleability.

- Use a more realistic velocity profile

The proposed model predicts the solids profile in the separator and is used in simulations to calculate the sludge blanket level and underflow concentration.
Figure 2.25 Solids concentration profile in the separator

After Stenstrom (11)
Solids loading 6.3 kg/m$^2$/hour (31 lbs/sq.foot/day)
Hydraulic loading 33.6 m$^3$/m$^2$/day (824 gal/sq.foot/day)

Solids loading 2.25 kg/m$^2$/hour (11 lbs/sq.foot/day)
Hydraulic loading 12 m$^3$/m$^2$/day (294 gal/sq.foot/day)

Figure 2.26 Solids concentration profile predicted by the model
2.5.3.3 Overall Separator Model

For this study, the purpose of the separator model is to allow the study of control strategies using computer simulations.

Clarification is not included in the overall model for the separator. If the purpose of the model was to expand the existing knowledge of separator dynamics it would be necessary that it predicts effluent suspended solids. However, the purpose of the model for this study is to establish control strategies for the process. Control strategies proposed herein for improving clarification are based on the research of Olsson (67). Olsson has shown that rapid increases in flow rate have a detrimental effect on clarification. Therefore, a control strategy is proposed in this study to reduce the hydraulic shocks to the separator. Although Olsson's model was not used in this study for actually predicting the effluent suspended solids concentration, the conclusions from Olsson's research did provide an objective function for control of clarification.

Effluent suspended solids concentration can only be predicted by the model used herein if solids in the effluent are a result of a failure of the separator to perform its thickening or storage functions. Such failures can occur due to separator overload. Overload was simulated using two flow rate inputs to the separator. Figure 2.27 shows two
Figure 2.27 Input flow rate patterns used in simulations
sinusoidal patterns used as input flow rate, one to simulate normal conditions (Figure 2.33a) and another with a pulse superimposed on this signal (Figure 2.33b) to simulate a sudden increase in load as might be caused by a storm. The underflow rate was maintained constant throughout the run.

Figure 2.28 shows the response of the solids profile to a sinusoidal input. The solids concentration is high in the lower part of the separator (below the feed point). During heavier loading, the sludge blanket rises somewhat but is still some distance from the top of the separator. The model predicts no solids in the upper portions of the separator.

Response to the pulse change in flow rate is shown in Figure 2.29. An increase of solids concentrations can be observed in the upper layers of the separator at the time the disturbance was introduced. This occurs because of the assumption in the model that the mass flux into a differential volume can not exceed the flux which the volume is capable of passing nor exceed the flux which the next higher differential volume is capable of transmitting. The sludge blanket rises to the top layer of the separator and as a result high concentrations of solids appear in the effluent.
Figure 2.28 Dynamics of solids profile in the separator for a 24 hour period. Flow pattern from Fig. 2.27 A used as input.
Figure 2.29 Dynamics of solids profile in the separator for a 24 hour period. Flow pattern from Fig. 2.27 B used as input.
2.6 Air Supply and Distribution System

The activated sludge process requires substantial amounts of energy with the air supply system being the single largest "energy user". Savings in energy used for aeration are thus essential for reducing operating costs. These savings may be obtained by controlling the air supply system.

A typical air supply system consists of several compressors, the air distribution system, and the diffusers. A schematic of the system modeled in this study is shown in Figure 2.30. The distribution of air flow to each aeration basin could not be controlled at Sagemont, but the modeled system was assumed to be capable of delivering the required amount of air to any aeration basin. Characteristic curves for the compressors used at Sagemont, as supplied by the manufacturer, were used in the model and are presented in Figure 2.31. There are four compressors operating in parallel and discharging into a main header.

The control system is designed so that the pressure in the main header is maintained at a constant value by regulating the output from the compressors. Air flow between the main header and each aeration basin is manipulated by the dissolved oxygen controller for each basin using valves between the main header and each basin. The total air flow demand is calculated as the summation of air flow rates to each basin.
Figure 2.31 Performance curves for compressors
At Sagemont, the compressors were controlled manually. A compressor control diagram (Figure 2.32) is used by the model to determine the on-off status and flow rate for each compressor for a given total air flow rate. This control strategy was adopted from the report on DO control by Flanagan et al. (69). After determining the flow rate through each blower from the blower control diagram, the energy utilization for each blower is calculated from the characteristic curve (Figure 2.31). The horsepower curve was approximated by a quadratic equation:

\[ HPOWER = D1 + D2 \text{ARFLOW}^2 \]  \hspace{1cm} (2.55)

**HPOWER** = horsepower  
**ARFLOW** = air flow rate  
**D1, D2** = parameters

Parameters A and B were estimated using least squares programs from the IMSL subroutine library.

Diffuser design has a strong effect on the performance of the air supply system. For example, fine bubble diffusers are more efficient than coarse bubble diffusers because the interface area between the bubbles and the liquid is larger. In the model, the effects of diffusers are incorporated into the value of the oxygen transfer coefficient. Goto (57) experimentally determined the oxygen transfer rate for the Sagemont plant and his result was adopted here. The value of the coefficient is therefore specific to the Sagemont
Figure 2.32 Schematic of control strategy for compressors

Four compressors operating in parallel

After Flanagan et al. (69)
plant and reflects the diffusers and conditions (depth of diffusors, tank geometry) at that particular plant. In order to estimate \( K \), Goto applied a linear least squares method to the following equation:

\[
\frac{dDO}{dt} = -K \cdot DO(t) + (K \cdot DOS - VTOUR)
\]  

(2.56)

where:

\( \frac{dDO}{dt} \) = time derivative of the DO, mg/L/hour  
DO\( (t) \) = DO concentration at time \( t \), mg/L  
VTOUR = oxygen demand, mg/L/hour  
\( K \) = oxygen transfer coefficient, m3/mg  
DO\( S \) = DO saturation concentration, mg/L  

Data for parameter estimation were obtained in a series of off-line experiments in which step changes were introduced in the air flow rate to the basin. The data analysis was performed assuming that:

1) The reactor was a single, completely stirred tank reactor with constant volume  
2) The transfer coefficient is a function of air flow rate only

Total oxygen uptake rate and mixed liquor temperature were measured and found to be relatively constant during the experiment. The following equation was used to relate oxygen transfer to air flow rate:
\[ K = E_1 + E_2 \times AFR \]  \hspace{1cm} (2.57)

where:

\begin{itemize}
  \item \( E_1, E_2 \) = parameters
  \item \( AFR \) = air flow rate to basin, in acfm
  \item \( K \) = plant specific oxygen transfer coefficient, l/hour
\end{itemize}

The value of \( K \) is influenced by the diffusers and the geometry of the tank. Since the data obtained from Sagemont was used to estimate the parameters \( A \) and \( B \), \( K \) is specific to that particular plant.
3. PROCESS CONTROL

3.1 Approach to Control

This chapter examines each unit of the system from a control point of view. The mathematical model of the plant described in the previous chapter is used to develop and evaluate the control strategies. The controllers are superimposed on the model of the plant and the effects of control on energy utilization and plant performance are explored through computer simulations.

There are generally three goals one tries to achieve in plant operation. Ranked in order of priority, these are:

- Avoid gross process failure, such as the gross discharge of solids from the solids-liquid separator
- Meet the permit limits on effluent quality
- Minimize operating costs (energy, chemicals, manpower)

In this study, control has been organized with these goals in mind. The control strategies are designed to apply in normal, routinely encountered conditions that prevail most of the time. During this type of operation the objective is to minimize operating costs, and permit limits are treated as constraints. Occasional upset conditions, usually the result of a significant disturbance to the process or equipment failure, are treated separately and will be discussed later. A schematic of the overall control
structure is presented in Figure 3.1.

During an upset condition the objective is to remedy the critical situation and prevent gross process failure. If an upset condition is detected special procedures are activated to handle the disturbance and return the plant to normal conditions.

The approach to the control problems studied in this research was strongly influenced by the work of Beck (22,23) and Joyce and Ortman et al. (5,28), as well as the literature on artificial intelligence (AI) and expert systems. Some of the AI concepts, and ideas concerning expert systems, will be presented in this chapter, along with possible implications of both for wastewater treatment control.

3.1.1 Expert Systems

There are two basic goals in artificial intelligence research:

1) Make machines smarter

2) Obtain a better understanding of human intelligence

The first goal, a more pragmatical one, has resulted in the development of expert systems. Expert systems are algorithms that solve problems using an approach similar to the one used by human experts. These programs are also referred to as rule based systems. Our goal, more
Figure 3.1 Overall control structure
specifically, is to establish techniques for improved operation of treatment plants.

What is the nature of expertise? Davis (70) defines expertise as the ability to perform the following tasks:

- Solve the problem
- Explain the result
- Learn
- Restructure knowledge
- Break rules
- Determine relevance
- Degrade gracefully

The above mentioned tasks are all a portion of the process an expert goes through when confronted with a problem. An expert is defined as a person who is knowledgable about a certain problem. A computer program capable of applying the same expertise, and obtaining a correct solution when presented with a problem, is called an expert system. Treatment of certain problems is more likely to be successful - there is definitively a category of problems that an expert system would be unlikely to solve. Davis (70) defines several characteristics of a "good" problem:

- There are experts
Experts are provably better than amateurs

Task takes an expert sufficient time

Skill is routinely taught

Task domain has high payoff

Task requires no common sense

The commercial success of several expert systems, such as XCON, MYCIN, and DIPMETER ADVISOR has created a strong interest in expert systems in industry. XCON (short for Expert Configurer) is a program developed by McDermot (71), and is extensively used by Digital Equipment Corporation to configure their VAX computer system orders. MYCIN was developed at Stanford University (72), and is used for infectious disease diagnosis and therapy selection.

AI staff at the Schlumberger company produced DIPMETER ADVISOR (73) to serve as a tool for analyzing dipmeter data. A dipmeter is an instrument used in the oil well drilling industry which contains several sensors. It is lowered into the bore hole in order to collect data which is later plotted on logs and interpreted by experts since analysis of these data is a complex task. DIPMETER ADVISOR was successful in capturing this expertise and proved to be a useful tool.

All of the above mentioned systems are rule-based programs, containing knowledge obtained from recognized experts on
specific problems.

3.1.2 Fuzzy Sets

Another development with substantial potential for control of the activated sludge process comes from fuzzy set theory, first established in a classical paper by Zadeh (74). By definition, a fuzzy set is a class of objects with a continuum of grades of membership. There is frequently ambiguity about the status of a member of a certain class — a group of tall men or good saxophone players, for example. In ordinary set theory, a member either belongs to the set or not. In fuzzy set theory each member of the set has a number associated with it indicating a degree of membership to a fuzzy set (between 0 and 1). Relationship between the population and the degree of membership to the set is called a membership function. Ordinary sets, with a discrete (0 and 1), rather than continuous membership function, fail to describe most of the situations usually encountered in real-life situations. Fuzzy set theory provides a technique for handling qualitative information in a rigorous fashion. It establishes ways to deal with linguistic information much closer to the way people tend to think.

The application of fuzzy sets in control is best suited for processes with:

- Low signal-to-noise ratio
- Nonlinear behavior
- Strong interactions (coupling) between process units
- A significant amount of a priori information about the process is available only in qualitative form

The activated sludge process has all of the above characteristics. The problem of estimating the process state from noisy measurements has been handled in the past by using the Kalman filter (75), among other tools. A survey of methods for model identification and parameter estimation can be found in an article by Astrom and Eykhoff (76). Tong (24) reviews the applications of fuzzy sets to control, and lists several examples where fuzzy control performed better than conventional control. In an IIASA (International Institute for Applied Systems Analysis) report (23), Beck and coworkers explore the applications of fuzzy control to the activated sludge process.

3.1.3 SCOUR Control

The specific oxygen uptake rate (SCOUR) represents the amount of oxygen utilized by microorganisms divided by the MLSS, and is used in this study to indicate the "activity" of the MLSS. Control of the SCOUR profile along the reactor is an important part of the control strategy proposed in this research. The SCOUR controller developed in this study and described later is essentially a rule-based system. It extracts information about the process from measurements and
reacts to perturbations from the desired state. Strictly speaking, fuzzy control is not directly used in this work in the sense that membership functions were not developed for any variable. In future versions of the control system presented here, fuzzy logic may be incorporated into the rules. The structure of the rules may therefore change, but it would still remain a rule-based system.
3.2 Pump station

Three objectives in the control of the pump stations are discussed in this study:

- Transporting the wastewater from the wet well and into the biological reactor at minimal cost
- Minimizing hydraulic disturbances to the solids-liquid separator
- Distributing the influent load to the biological reactor in the most favorable way from the standpoint of the biological reactions

In this chapter, control strategies for achieving objectives 1 and 2 will be examined. The second objective will also be discussed in section 3.3 with regard to solids-liquid separation. The third objective can be treated independently of the pump station, and will be discussed in the section on the biological reactor.
3.2.1 Model of Pump Station Control System

Three different pump control systems were modeled and examined. All data for the models were obtained from the pump station at the Sagemont treatment plant. Detailed information concerning the pump station and control systems used for the purpose of this work can be found in engineering reports by Andrews (8-17), and literature from the pump control system manufacturer (46). A complete listing of the computer program for simulation of the pump station and the control systems is given in Appendix 8.2.

The three control systems considered are:

- On/Off pumping
- Automatic throttling of the pump discharge
- Variable speed pumping

Computer models of different pump control systems were developed and superimposed on the pump station model presented in section 2.3.

Figure 3.2 presents a general flow chart for the overall model. One minute averages for the effluent flow rate were used in the simulations. These data were stored in a file and used as an input to the model. Every time interval (one minute), the procedure shown in the flow chart is repeated. During the first step indicated on the flow chart influent
Figure 3.2 Overall flow chart for pump station control model

START

Step 1
Obtain influent flow rate data
Qin

Step 2
Calculate flow out of the pump station
Qout

Step 3
Calculate new wet well level from the mass balance
flow rate to the wet well is read from the data file.

In step 2, shown in detail in Figure 3.3, the desired flow rate for the pumps is calculated according to the wet well level and the control strategy for each pump. Speed of the variable speed pump and the position of the automatic throttling valve were determined from the diagrams shown in Figure 3.4 and Figure 3.5. Pump curves were represented by a third order polynomial (eq. 2.6) with the coefficients for the polynomial being calculated from the pump speed. The system head curve was represented by a second order polynomial (eq. 2.5) with the coefficients being calculated from the wet well level. The system head for the pump controlled by the throttling valve depended also on the throttling valve position so this is included in calculation of polynomial coefficients for the system head.

Figures 3.4 and 3.5 were generated from the model of the control system at Sagemont and follow specifications provided by the control system manufacturer (46). These diagrams were considered typical of conventional pump controllers.

In step 3 (Figure 3.2), a new wet well level is calculated from the mass balance around the wet well. The same procedure is then repeated for the next time interval.
Figure 3.3 Flow chart for a single pump controller
from Maintenance Manual for Louis Allis Lancer 44XLP
Simplex Drive System for Worthington 6-860640 125 HP pump

Figure 3.4 Valve control diagram for automatic throttling of the pump discharge
from Maintenance Manual for Louis Allis Lancer 44XLP
Simplex Drive System for Worthington 6-860640 125 HP pump

Figure 3.5 Speed control diagram for variable speed pumping
3.2.2 Reducing Hydraulic Shocks

There is evidence in the literature (Olsson & Chapman (67), Crosby (68)) to support the hypothesis that turbulence caused by rapid changes of flow rate into the solids-liquid separator causes an increase in effluent suspended solids concentration. It would therefore seem desirable to manipulate the flow rate into the solids-liquid separator with the objective of improving clarification. This chapter examines the possibility of controlling the flow rate from the pump station with the objective of creating optimal flow patterns for the separator.

Figure 3.6 shows a plot of flow rate vs. time during a single day at the Sagemont plant. The flow rate is measured at the plant effluent. Since some flow comes from remote pumping stations and does not go through the wet well and there is some dampening of flow in the plant, it is not identical to the flow into the wet well. However, the influent flow was not measured so the effluent flow measurement had to serve as an approximation.

Figure 3.7 shows the results of a simulation of the variable speed pump station control system. The flow rates shown in Figure 3.6 were used to simulate input into the wet well, and the computer model described in section 4.1.1 was used to simulate the operation of the pump control system.

It can be seen that the flow rate coming out of the pump
Figure 3.7  Output from the pump station using variable speed pumping - simulated results.
station is almost identical to the flow rate at the input to the wet well. High frequency components in the influent flow rate are also present in the output, and sharp changes are not dampened. This result can be easily explained by the fact that the volume of the wet well is rather small (106 cubic meters between the low and high levels). A simple mass balance for the wet well shows that due to its small size it is not capable of substantially changing the character of the flow rate. A material balance on the water in the wet well gives:

\[ A \times \frac{dh}{dt} = Q_i - Q_o \]  

(3.1)

where:

- \( A \) = cross-sectional area of the wet well
- \( h \) = level
- \( Q_i \) = input flow
- \( Q_o \) = outlet flow
- \( t \) = time

Where, \( h \) is constrained by:

\[ h_{\text{max}} > h > h_{\text{min}} \]

at Sagemont \( h_{\text{max}}=11 \) feet, \( h_{\text{min}}=4.2 \) feet

At time \( t \), it is assumed that a flow imbalance, \( B \), exists. \( B \) is given by:

\[ B = Q_i(t_0) - Q_o(t_0) \]  

(3.2)
In order to smooth the flow coming out of the wet well, the variation in $Q_0$ has to be minimized. McDonald et al. (77) provide a solution to this problem:

$$Q_0(t) = \frac{B**2(t-t_0)}{2A(h_{\text{max}}-h(t_0))} + Q_0(t_0) \quad (3.3)$$

It is apparent from the equation above that the cross sectional area of the wet well directly influences the ability of the wet well to reduce variation in the flow rate. For the Sagemont plant, the wet well contained 45 cubic meters of water per meter of depth. It is not possible to significantly change the shape of the flow rate without some capacitance - a reservoir in which the level is allowed to fluctuate and therefore provide dampening capacity.

There are two possible volumes that may be used for the purpose of dampening the changes in flow rate before and after the wet well, namely the sewer network and the biological reactor. Computer control of the sewer network is used in many cities - Minneapolis (78), Detroit (79), Cleveland (80) and Seattle (81) are some examples. The interaction between the operation of a treatment plant and the sewer system has been examined by Nelson et al. (82), among others. The use of the sewer network to provide dampening capacity was not examined in this research, since it fell outside the boundaries of the system being examined, as shown in Figure 1.1. Also, storage of sewage in sewers may not be feasible in hot climates such as Houston's
because of possible corrosion and odor problems.

The use of a variable volume biological reactor is proposed here in order to provide dampening capacity. A small (0.3 meters) variation of level in the biological reactor can provide a much greater volume for dampening than that available in the wet well. In the case of the Sagemont plant, the cross-sectional area of the biological reactor is approximately 30 times greater than the cross-sectional area of the wet well.

A change of level in the biological reactor affects the aeration system by changing the depth of immersion of the air diffusers. The pressure set point (in the main air header) for the compressor controller needs to be selected on the basis of the highest level in the aeration tank. A decrease in pressure would cause an increase in air flow rate into the basin, therefore there is no risk of encountering surge. Although the level in the biological reactor changes only 0.3 meters under the proposed control strategy the interactions between the aeration system and the liquid level in the reactor should be taken into account in the design of the aeration control strategies. For example, the pressure set point in the main air header could be manipulated according to the level in the tank, allowing a decrease in the pressure set point with a decrease in level.

If flow rate is controlled between the biological reactor
and the separator, the pump station and the separator are decoupled. In most treatment plants several separators operate in parallel and there is a need for control of the distribution of flow between individual separators so that flow rates are the same to each separator. Control of the flow rate between the biological reactor and the separator would thus provide the dual benefits of proper flow distribution and reduction of hydraulic shocks.

Strategies for control of flow rate between the biological reactor and the separator will be examined in Chapter 4.
3.2.3 Minimization of Energy Utilization

The design of wastewater pumping stations and the selection of pump control strategies is a difficult dynamic problem because of the wide variations in flow rate which must be accommodated, the limited storage capacity available in the wet well, and the need for high reliability. The objective function adopted here for controlling the pump station is to minimize the energy requirement for pumping of influent wastewater into the process. However, there are examples in the literature with different objective functions for control of the pump station. Shioja et al. (38), for example, minimized the number of starts and stops in order to prolong pump and motor life and reduce maintenance requirements.

Data obtained from the Sagemont plant will be used, in addition to a mathematical model of the pump station (section 2.3), for examination of the effect of different control strategies on energy requirements. The observations and simulations presented are therefore limited in applicability to low-friction head, high-static head pump stations such as the one at Sagemont.

Two control strategies will be compared on the basis of energy utilization - on/off control and variable speed pumping. Conclusions will be limited to single pump operation, and pump stations with characteristics similar to the one from which data were obtained. Simulations will be
performed using the mathematical model described earlier in the text.

3.2.3.1 On/Off vs. Variable Speed Pumping

Power utilized by the variable speed pump was calculated from the following expression:

\[ P_i = \frac{(g_a F H_t)}{Eff} \]  \hspace{1cm} (3.4)

where:

- \( P_i \) = Power input (L Force/T)
- \( g_a \) = specific weight of water (M/L**3)
- \( F \) = flow rate (L**3/T)
- \( H_t \) = total dynamic head (Force/L**2)
- \( Eff \) = overall efficiency = \( E_p \times E_d \times E_m \)
- \( E_p \) = pump efficiency
- \( E_d \) = variable speed drive efficiency
- \( E_m \) = motor efficiency

The efficiency curve provided by the pump manufacturer was used in expression 3.4. Figure 3.8 shows the calculated power required for pumping as a function of flow rate at two wet well levels. The power required was monitored by measuring power input to the pump motors. The results of these measurements agree with the theoretical results, and are presented in Figure 3.9.

If a single variable speed pump is compared with a single
Figure 3.8 Calculated power required for variable speed pumping
Figure 3.9 Measured power requirements for variable speed pumping
On/Off pump, within the constraints given earlier as to the type of pump station, the On/Off pump performs more efficiently than the variable speed pump. In order to meet the peaks in the demand for pumping, the variable speed pump has to be large, and performs poorly during periods of lower flow rate. The main reason is that the pump efficiency is very low at low flow rates, as can be seen from the pump curve (Figure 2.11).

Operation of the station at the highest possible wet well level reduces energy utilization since the pump is working against a smaller total dynamic head. The effect of the level in the wet well on energy utilization is presented in Figure 3.10.

The control strategy that keeps the wet well level as high as possible at all times reduces somewhat the cost of pumping. In cases where the variable speed pumps are used, it is possible to use the predictions of influent flow rate and move the set points for the levels in the wet well higher, if sudden increases in flow rate are not expected, and lower during the periods of higher loads.

Control of the variable speed pump was simulated for a 24 hour period. Several simulations were made using the same influent flow rate, but with the levels in the wet well used for on/off control of the pump being changed for each simulation. For the same influent flow rate, energy utilization for the on-off pump was 345 kWhr, which is
Start and stop levels in the wet well were raised to minimize static head and increase the flow rate through the pump.

Figure 3.10 Effects of level set points on energy utilization.
substantially less that for the variable speed pump (900-990 kWhrs). These result is in agreement with energy utilization observed at Sagemont. Vitasovic et al. (83) report that on an average dry day energy utilization was 201 kWhrs and 904 kWhrs for On/Off and variable speed pumping, respectively.

The variable speed pump performs better at higher flow rates. Table 3.1 shows the results of simulations with different influent flow rates. Sinusoidal signals, as defined in Table 3.1, were used for influent flow rate.

The predictions of the simulations are confirmed by the results obtained at Sagemont.
Table 3.1
Energy Utilization for Different Flow Rates

<table>
<thead>
<tr>
<th>Flow in L/S</th>
<th>Amplitude</th>
<th>Power Used in kWhr</th>
<th>Var. Speed</th>
<th>On/Off</th>
</tr>
</thead>
<tbody>
<tr>
<td>110</td>
<td>110</td>
<td>787.12</td>
<td>248.1</td>
<td></td>
</tr>
<tr>
<td>153</td>
<td>110</td>
<td>962.54</td>
<td>345.1</td>
<td></td>
</tr>
<tr>
<td>200</td>
<td>88</td>
<td>1027.8</td>
<td>441.43</td>
<td></td>
</tr>
</tbody>
</table>
3.3 Biological Reactor

The activated sludge process is capable of producing a high quality effluent but it is also an energy intensive process. The energy requirement for supplying air to the biological reactor is by far the highest energy requirement for most wastewater treatment plants. Air has to be supplied to the biological reactor in order to keep the MLSS in suspension, and to provide enough oxygen to meet the demand created by biological activity. Since it is the largest "energy user", it is the most promising area for application of process control to reduce operating costs. Most research in control of the activated sludge process is there centered around control of aeration and many studies have been reported in the literature. Some of the more typical studies are reviewed in the following section.

A description of the control strategy proposed in this research is presented in section 3.2.2. Problems concerning process state definition and estimation will be discussed and controller design will be addressed.
3.3.1 Literature Review

One of the early reports on operation and control of the activated sludge process, published in 1942 (84), considers oxygen uptake rate as one of the most important operational parameters and recognize dissolved oxygen concentration as a constraint on aerobic biological activity. For example, in Chicago dissolved oxygen concentrations were manually measured every four hours (Winkler method) and the results used to manually control the air supply to the reactor.

The 1942 report also discusses such concepts as "sludge activity" and the "biochemical condition" of sludge. It gives several proposed definitions for these parameters. Ruchhoft (85) defined activated sludge "condition" as its capacity to oxidize organics. He observed that the ability of sludge to oxidize organic material depended on the oxygen uptake rate (OUR) of the sludge and stated that the OUR of mixed sludge liquor is an important criterion of the biochemical condition of a sludge.

Bloodgood, one of the coauthors of the report, proposed the use of SCOUR to estimate the status of the biological reactions. He used respirometers for control of a full scale wastewater treatment plant and suggested that high SCOUR values could result in sludge that does not concentrate well.

The report concludes that sludge condition, as described by OUR, is an important operating variable.
In his 1948 article, Torpey (86) introduced the useful concept of step feed. He shifted the influent from the head of the plant and introduced it at several points along the length of reactor, thus creating a range of contacting patterns between the sewage and mixed liquor. Step feed thus serves as a manipulatable variable for improving the efficiency and stability of plant operation. Step feed has since been shown by many researchers to be a valuable tool for process control.

Sorensen (87) examined the use of step feed control in a pilot plant in Denmark. Different contacting patterns were simulated and a control strategy based on OUR was examined. He sites effluent quality control, energy savings, and reduced excess sludge production as possible benefits of step feed control. In later work, Sorensen (88) examined the use of step feed to maintain a constant food to mass ratio (F/M).

Dissolved oxygen concentration imposes a constraint on the performance of a biological reactor, since it should be maintained above the value where it limits the growth of aerobic organisms. Since the cost of aeration is a major operating cost in activated sludge, there are a number of instances (89,90,91) where the DO (usually measured near the outlet of the biological reactor) is controlled at a value that will allow adequate aerobic biological activity, but avoid wasting of energy. A single measurement of dissolved
oxygen at one point in a reactor does not, however, contain sufficient information on the performance of the biological reactor. Dissolved oxygen concentration is a constraint that should not be violated, but if the dissolved oxygen in a reactor is at the setpoint it does not necessarily imply that the reactor is in a proper operational state. Since DO is a reliable physical measurement obtainable on-line and can be used as a signal to control air flow rate, it has been a popular choice for use in control of the activated sludge process.

Olsson and Andrews (25) have suggested that information obtained from the DO concentration profile along the reactor can be used to estimate the state of the process. They identified the following features of the profile as significant:

- the slope near the outlet
- the concentration at the outlet
- maximum slope
- the position of the maximum slope
- the "convexity" at the outlet (second derivative)

The slope near the outlet is related to the substrate concentration and indicates the completion of the biological reaction. If the reaction is near completion, organism decay becomes the dominant reaction and the profile curve approaches the horizontal. The position of the maximum slope indicates the loading to the reactor. During higher
loadings the position of maximum slope is pushed towards the end of the reactor. The maximum slope of the profile reflects the organism growth rate in the reactor. The dissolved oxygen concentration at the outlet and the second derivative of the profile are other indicators of the completion of the biological reaction. However, the author is not aware of any automatic control applications utilizing the information obtained from the profile as described by Olsson and Andrews.

Takamatsu et al. (92) recognized a need to estimate the state of the process and proposed OUR as a process variable. He utilized OUR for control of a laboratory scale unit with recycle flow as the manipulative variable. A proportional-integral (PI) controller was used with the coefficients for the controller being obtained from the results of step tests.

An interesting application of step feed control has been proposed by Yust et al. (93). In this research, step feed was used to control SCOUR in the last stage of a reactor. SCOUR indicates the degree of completion of the biological reaction and the objective was to keep it constant. The set point for SCOUR was described as being plant specific, in that it depends on the plant design, type of waste etc. From the setpoint for SCOUR, the desired concentration of sludge in the last portion of the reactor was calculated as follows:
\[ X_{\text{des}} = \frac{\text{OUR}3}{\text{SCOURsp}} \]  

where:

- \( X_{\text{des}} \) = desired MLSS concentration in last reactor
- \( \text{OUR}3 \) = OUR in last reactor
- \( \text{SCOURsp} \) = SCOUR set point

The required contacting pattern was then calculated from simplified mass balance equations in which biological growth was assumed negligible. This is a reasonable assumption for short periods of time and especially when sludge wasting is properly controlled.

Stenstrom and Andrews (94), using a dynamic model and computer simulations, examined control strategies based on SCOUR. They investigated the minimization of variability of SCOUR in time using two types of control - step feed and recycle rate control. The results of their simulations indicated a significant decrease in SCOUR variability over time, up to 86%. The use of an on-line flow predictor in control improved the performance of step feed control 48% as compared to control without the flow prediction term.

Beck et al. (22) proposed the use of rule-based systems and fuzzy logic for control of the process. They developed a set of rules for operation of the process and suggested the use of controller algorithms based on fuzzy logic. In order to facilitate the design of such controllers they proposed development of questionnaires to be answered by operating
engineers. The questionnaire would be a communication tool between experienced plant operating engineers and control engineers. Operating rules would then be developed to formalize this qualitative information.

An innovative control strategy based on a study conducted at a full scale plant in Hillsboro, Oregon is presented in reports by Zickefoose, Joyce and Ortman (5,28). The state of the process was determined from field measurements, and a rule-based control system was designed to keep the process in the optimal operational state. The state was assumed to be described by the following parameters:

- food to mass ratio (F/M)
- respiration rate (RR)
- sludge settling volume (5 min.) (SSV)
- dissolved oxygen concentration (DO)

Both food (influent substrate concentration) and mass (microorganism concentration) were determined by total organic carbon (TOC) measurements. Respiration rate was measured off line using a DO meter.

Air flow rate was controlled with a feedback controller using dissolved oxygen as a process variable and was independent of the "main" control strategy. The overall control strategy used three parameters - SSV, RR, and F/M. Three regions were established for each parameter - low, optimum and high. The state of the system was determined from the state of the parameters, and since there were three
parameters with three possible states, 27 different states of the system were possible. The concept was visually represented with a three dimensional, 27-element matrix as shown in Figure 3.11.

The optimal state was defined as the one in which all three parameters are in their optimal ranges (the center of the matrix). For each of the remaining 26 operational states defined as not optimal, rules were established from operating experience that described the control actions needed to return the process to the optimal operational state.

Recycle sludge flow rate, waste sludge flow rate and the step feed point (related to sludge conditioning time by the authors) were manipulated according to the operational rules. For example, if the sludge settling volume is low, with respiration rate and F/M still in the optimum range, the process operates in the state indicated by element 5 in the operating condition matrix. The rule for element 5 indicates that the recycle flow rate has to be increased. This rule-based control strategy resulted in significant improvement in operation over the uncontrolled case.

In concept, this control strategy is quite similar to the idea of rule based control system described by Beck et al. (22).
Figure 3.11 Operational condition matrix

After Joyce et al. (34)
3.3.2 Proposed Control Strategy

The control strategy proposed here was developed using a mathematical model to simulate the behaviour of the plant. Outputs of the model were used by control algorithms so that real-time control could be simulated. All simulations were performed for the conventional activated sludge process with step feed capabilities and a mean cell residence time of 10 days so that nitrification could be accomplished.

All the information used by the controllers would be available on line in an instrumented activated sludge plant, either directly from sensors or estimated from the measurements obtained by the sensors. A possible exception is the value of SCOUR in each reactor. The controllers presented here use the values of SCOUR calculated from the model, and do not attempt to estimate SCOUR from the other measurements. In a real situation, SCOUR would have to be estimated on-line from the measurements. For a survey of methods for on-line SCOUR estimation, the reader is referred to Goto (57).

The general problem of controlling the process can be divided into three parts:

- Establish a technique for determining process performance. It is necessary to have a method that will provide an on-line estimate of the process state.

- Define the desired state of the process. In other
words, establish the theoretical basis for the
definition of the objective function.

- Design the controller. Determine what control action
will most effectively change the process state so as to
minimize perturbations from the optimal process state.

Answers will be proposed to these problems. First, the
state of the process will be defined. After this,
parameters are proposed that can be used to estimate the
state on-line.

There are three main goals that a well operated biological
reactor should achieve:

- High degree of BOD removal

- Produce solids with good settling properties

- Achieve 1 and 2 with minimum energy utilization

In order to estimate the state of the biological reactor, it
is therefore important to establish means for monitoring the
process and measure the achievement of these goals.

One way of measuring BOD removal is to sample the plant
influent and effluent and perform standard BOD tests.
However, five days are required to obtain the results so
this is not very useful for control purposes. There are
faster measurements that can be used such as TOD, COD or
TOC, but some of the instruments are quite complex and the
measurements do not always correlate well with BOD.

According to the structured model for the biological reactor, at low substrate concentrations the dominant biological reactions would be utilization of stored mass and organism decay. The amount of oxygen being consumed would therefore be significantly less in the last reactor than in the preceding reactors where substrate concentrations are higher and oxygen is consumed in reactions for substrate utilization. In this study, the value of SCOUR in the last reactor will be used as an indicator of the degree of completion of the reactions. It may also be indicative of the settling characteristics of the sludge.

The second goal is to create sludge with good clarification and thickening properties, since this is very important both for producing a high quality effluent and stable operation of the solids-liquid separator. The conditions in the biological reactor affect the properties of the sludge and, therefore, operation of the separator. By controlling conditions in the biological reactor, this interaction can be used to improve plant operation.

In their article on activated sludge bulking Sezgin, et al. (95) discussed several conditions that may lead to preferential growth of filamentous microorganisms and thus development of sludge with poor settling characteristics. They defined three main types of sludge:
Case 1: Too many filamentous organisms in the sludge resulting in a slowly settling, poorly compacting sludge.

Case 2: A sufficient number of filamentous organisms to provide strong floc and sludge that settles and compacts well.

Case 3: Insufficient filamentous organisms leading to easy break-up of the floc and the creation of pin floc.

They further propose that growth conditions strongly affect the type of sludge produced in the biological reactor. For the case where oxygen is the limiting factor in growth of both filamentous and zoogloeaal organisms, they propose that, at low dissolved oxygen concentrations, filamentous organisms grow more rapidly than zoogloeaal. The opposite is assumed to be true at high dissolved oxygen concentrations. The critical DO concentration was identified as the concentration at which both groups of organisms had the same specific growth rate. This would imply that if oxygen is at the critical concentration, conditions would most likely be favorable for the development of Case 2 floc. Above the critical concentration, zoogloeaal organisms have a greater growth rate and conditions are most likely to result in production of Case 3 sludge. A region below the critical concentration would provide conditions most likely to result in greater growth rates for filamentous organisms and the development of Case 1 floc.
In measuring dissolved oxygen, the concentration measured by the DO probe is that in the bulk solution and not within the floc. There is a DO profile within the floc, starting from the concentration in the bulk solution and decreasing towards the center of the floc due to oxygen uptake by the microorganisms, as shown in Figure 3.12. Depletion of DO between the surface of the floc and its center is related to the value of SCOUR.

Two important conclusions may be drawn from the above:

- The dissolved oxygen concentration measured in the bulk fluid does not itself contain sufficient information on conditions within the floc
- The oxygen uptake rate influences the dissolved oxygen profile in the floc

According to Sezgin, et al. (95), low DO concentrations in the center of the floc may result in preferential growth of filamentous organisms and development of Case 1 floc.

It should be emphasized that there are many other factors, in addition to dissolved oxygen concentration within the floc, that are significant for the development of non-bulking activated sludge. The composition of the wastewater, for example, has a pronounced effect on bulking. Wagner (96), reporting on experiences with sludge bulking in Germany, states that microscopic inspection of about 3500 sludge samples from 315 activated sludge plants indicated
Figure 3.12 DO profile within the floc after Sezgin et al. (95)
that more than 45% of the plants had extensive filamentous populations in the sludge. Wastewater with an unbalanced nutrient composition, such as effluents from the paper industry, milk and fruit processing, and breweries, are more likely to form bulking sludge.

The degree of longitudinal mixing and the substrate concentration gradient along the reactor have been cited by Eikelboom (97), Tomlison (98), and Grau et al. (99) as important factors affecting sludge settling properties. They proposed the use of plug-flow reactors, claiming that the low substrate concentration gradients in complete mixing systems create conditions for preferential growth of filamentous microorganisms.

Eikelboom (97) examined the problem of bulking at pilot-scale and proposed a method to control bulking due to low substrate concentrations. He introduced a biological contactor with a very short (10-20 minutes) hydraulic residence time ahead of either a complete mixing reactor or a plug flow reactor. This type of operation is based on the premise that their large surface area gives filamentous organisms a competitive edge in low substrate environments. Separating the exogenous and endogenous growth phase, achieved by the use of a contactor, or in a plug-flow regime, would create favorable conditions for the development of non-filamentous organisms. Eikelboom also observed that sufficient time has to be provided for the
endogenous phase in order to achieve high rate biosorption during the exogenous phase.

One key measurement (DO concentration) and one key parameter (SCOUR) will be used to estimate the conditions of the floc. The DO measurement indicates the DO concentration in the bulk solution, and the value of SCOUR gives an indication of the DO profile within the floc. It is the goal of the control strategy for the biological reactor to operate so that favorable conditions are created for the development of Case 2 floc. Results of the study by Sezgin, et al. (95) were used to define favorable process conditions (on the micro level), in terms of parameters obtainable on-line.

The assumption made is that oxygen is the limiting factor in growth. An attempt is made in this study to control the growth conditions. The objective is to create conditions most likely to produce sludge with good settling characteristics. Since SCOUR is an excellent dynamic indicator of growth, it is used as a process variable.

The third goal is to minimize energy consumption, provided that the plant effluent is meeting permit requirements and operating conditions are favorable for the development of a good settling sludge. Energy is saved through implementation of dissolved oxygen control. The dissolved oxygen concentration is maintained as low as possible without negatively affecting the biological reactions. This provides the largest possible difference between the actual
and the saturation oxygen concentration thus maximizing the driving force for transport of oxygen from the gas phase to the liquid phase.

Dissolved oxygen control is limited by two constraints. The upper limit is set by the maximum capacity of the aeration equipment and the lower limit is set by the amount of air necessary to provide adequate mixing.
3.3.2.1 Overall strategy

Control of the biological reactor is achieved by three controllers that run concurrently. These are:

- MLSS controller
- Dissolved oxygen controller
- SCOUR controller

Schematics for each of the mentioned controllers are presented in Figures 3.13 through 3.15.

Two different types of MLSS controllers were simulated. Both used waste sludge flow rate as the manipulative variable with sludge being only wasted from the last of four reactors in series. However, each used different process variables as signals for control. The first type of MLSS controller used information on the suspended solids concentration in each reactor to calculate the approximate mass of solids in all reactors. It is assumed that the mass of solids in the solids-liquid separator is negligible or remains constant. Wasting rate was then controlled to keep the total mass of solids in the system constant. The second type of controller reduces the number of measurements required by using only the suspended solids concentration in the last reactor as a process variable. A PI controller was used to determine the wasting rate.

The second approach was found to be more suitable for the
Figure 3.13  MLSS controller
purpose of this research since it was simpler and would be easier to implement.

The second part of the control system for the biological reactor is the dissolved oxygen controller. Dissolved oxygen concentration in each reactor is controlled with a standard PI controller. Air flow rates between the main air header and each of the biological reactors are manipulated, based on the perturbations from the dissolved oxygen concentration setpoints in each reactor. Control of the compressors does not depend directly on the dissolved oxygen concentrations, but rather on the pressure in the main air header with the objective of maintaining this pressure constant.

It is assumed that the valves, valve actuators and DO probes are ideal and their action is instantaneous and accurate. Control is initiated every 15 minutes in the model. In a real application, control could be initiated more frequently if needed.

Proper attention would have to be devoted to signal processing in actual applications. The effects of noise can and should be eliminated by filtering techniques. The dissolved oxygen concentration signal often displays high-frequency noise. At the Sagemont plant, one minute values were calculated as averages of ten six-second readings and this simple filtering produced a sufficiently noise-free signal.
Finally, the third control algorithm is the one that controls the SCOUR profile using step feed. Figure 3.16 shows the results of the simulation of 24 hours of operation of a biological reactor in an activated sludge plant. The surface represents SCOUR values in both time (hours 0-24) and space (reactors 1-4). The reactor is modeled as four completely mixed stirred tanks in series. No SCOUR control was exercised for this simulation, so it represents a base case for comparison. Both the MLSS controller and DO controller were active in this simulation. Input to the model was a sinusoidal signal in flow and concentration (in phase) representing the influent to the wastewater plant. The simulation conditions are summarized in Table 3.2.

The shape of the surface shown in Figure 3.16 is influenced by such values as the composition of the influent to the plant (fractions of slowly and rapidly biodegradable substrate), mean hydraulic retention time, MLSS concentration, and the values of kinetic parameters. The influence they exert on the SCOUR profile will be discussed later, with an "average" case being examined first.
MLSS = 4500 mg/L
Mean hydr. retention time = 5.55 hours
All influent fed into basin 1

Figure 3.16 SCOUR variation in space and time
No SCOUR control
Table 3.2
Summary of the Biological Reactor Simulation Conditions

<table>
<thead>
<tr>
<th></th>
<th>Mean</th>
<th>Amplitude</th>
<th>Period</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow Signal</td>
<td>450 m³/hr</td>
<td>25 m³/hr</td>
<td>24 hr</td>
</tr>
<tr>
<td>Concentration Signal</td>
<td>200 mg/L</td>
<td>100 mg/L</td>
<td>24 hr</td>
</tr>
<tr>
<td>Mean Hydraulic Residence Time:</td>
<td>5.55 hr</td>
<td></td>
<td></td>
</tr>
<tr>
<td>MLSS Set Point:</td>
<td>4500 mg/L</td>
<td></td>
<td></td>
</tr>
<tr>
<td>D.O. Set Point (each basin):</td>
<td>2 mg/L</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
It can be observed from the surface shown in Figure 3.16 that most of the variation and the highest values are both observed in reactor 1. The sinusoidal signal is significantly dampened in reactors 2-4.

Research by Jenkins and coworkers (95,100) suggests that DO depletion within the floc depends not only on the DO concentration in the bulk liquid, but also on the value of SCOUR. At higher values of SCOUR, DO depletion from the surface to the center of the floc is greater, creating conditions that may lead to poor settling properties. Thus, if SCOUR is controlled at a low value throughout the reactors, it should be possible to maintain the dissolved oxygen concentration at a lower level without increasing the risk of adversely affecting the settling properties.

It is be proposed here that SCOUR, and, therefore, organism growth, be controlled in both space and time. Stenstrom and Andrews (94) proposed control of SCOUR in time, taking an average in space. Control in both space and time represents a logical step forward. Palm et al. (95) suggested that an increase in removal rate at constant DO may mean depletion of oxygen inside the floc and consequent growth of filamentous organisms. This would imply that even with the DO controller performing rather well, high SCOUR values in the first reactor might mean that conditions are likely to result in low oxygen concentrations in the center of the floc. This would create an environment where growth of
filamentous organisms would be greater than that of zoogloal organisms, resulting in sludge with poor settling properties.

The objective of the SCOUR controller is to minimize variation in the surface shown in Figure 3.16, and to minimize the average value of SCOUR overall - the controller attempts to make the surface as flat and as low as possible. The assumption here is that the low and flat surface translates into the conditions for development of sludge with good settling properties. The value of SCOUR in the last reactor is a constraint, though, and should not exceed a value at which the biological reactions are incomplete.

The first method examined for SCOUR control was a simple discrete controller that routed the influent flow to reactor 2 if the set point for SCOUR was exceeded in reactor 1. Nothing was changed from the base case except that the SCOUR controller was activated. The results are shown in Figure 3.17. It is obvious that the surface is lower and flatter than in Figure 3.16. The range for SCOUR of 4 - 20 mg/L hour without control is now reduced to a range of 4 - 13.5 mg/L hour when control is exerted.

It would appear that rule-based controllers would be well suited for this type of control, so a controller was designed using a rule-based system to control SCOUR. The proposed control strategy controls the process in a fashion similar that of an operator but decisions are made
MLSS = 4500 mg/L

DO set point = 2.0 mg/L in each reactor

Mean hydr. retention time = 5.55 hours

Figure 3.17: SCOUR variation in space and time

Simple discrete controller
automatically. Since it simulates the "expert" actually running the plant, this may be referred to as expert control. In order to facilitate the design of the controller, a program was developed to generate a surface similar to the one obtained in the base case for the SCOUR profile in time and space. Since the purpose of the controller is to reduce the variation in the surface using one manipulative variable a complex biological model was replaced with a greatly simplified input-output relationship. This procedure simplified the controller design and reduced simulation costs. SCOUR in each reactor was assumed to be only a function of influent flow coming into that reactor and the influent flow coming into the reactor preceding it. The equation used for SCOUR was:

\[ \text{SCOUR}_n = A + B \times \text{Fn-1} + C \times \text{Fn} \]  

(3.6)

where:

\( \text{SCOUR}_n = \) SCOUR in reactor \( n \)

\( \text{Fn} = \) influent flow rate to reactor \( n \)

\( \text{Fn-1} = \) influent flow rate to reactor \( n-1 \)

\( A, B, C = \) constants

The surface generated when no SCOUR control is exercised is shown in Figure 3.18.
Figure 3.18 "Artificial" surface generated to represent SCOUR variation in space and time. This surface was used to facilitate controller design.
While the surface presented in Figure 3.16 was obtained as a result from a complete model for the activated sludge process, this surface (presented in Fig. 3.18) is generated by equation 3.6. using the same input signals for the influent flow rate.

The controller was applied to the "artificial" surface during the development of the controller. After satisfactory results were obtained the same algorithm was applied to simulate the controller using the complete model of the process.
3.3.2.2 Description of the Expert SCOUR Controller

The SCOUR controller uses values of SCOUR for each of the four reactors as input and calculates the desired contacting pattern for the next control period. It is important to note that the control period is defined as the interval in which the controller manipulates the feed point and is not equal to the sampling period. A control period of 15 minutes was used for the simulations presented. The sampling period would normally be less than this assuming data acquisition by computer. The objective is to minimize the offset from the setpoint in each reactor.

A setpoint (or "ceiling") is established for each reactor. It represents the largest value of SCOUR allowed for the next time interval for each reactor. It is necessary to initialize the controller by setting the initial ceilings for each reactor rather low, approximately at a value that would be observed in the last reactor when 100% of the flow is introduced into the first reactor.

As the influent flow changes, one of sixteen possible situations summarized in Table 3.3 will be observed. These values arise from a comparison of the values of SCOUR in each reactor with the ceilings set for each reactor. The value of SCOUR is compared with the ceiling in each reactor, and the results are summarized in a single number by a flag. This flag, with a value between 0 and 15, indicates roughly what state the process is in with regard to the SCOUR
<table>
<thead>
<tr>
<th>The value of the flag</th>
<th>Reactor 1 (LSB)</th>
<th>Reactor 2 (bit 2)</th>
<th>Reactor 3 (bit 3)</th>
<th>Reactor 4 (bit 4)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>2</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>3</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>4</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
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<td>1</td>
<td>1</td>
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<td>6</td>
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<td>1</td>
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<tr>
<td>8</td>
<td>1</td>
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<td>1</td>
<td>1</td>
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<tr>
<td>9</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
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<tr>
<td>10</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
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<td>13</td>
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</tr>
<tr>
<td>14</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td>15</td>
<td>1</td>
<td>1</td>
<td>1</td>
<td>1</td>
</tr>
</tbody>
</table>

Table 3.3 Possible states of the SOCUOR profile
profile. The flag is represented by four (binary) bits, with the least significant bit representing reactor 1 and the most significant bit representing reactor 4. The bits are set if the value of SCOUR in a given reactor exceeds the ceiling and are cleared if that is not the case.

An example of a rule used is:

\[
\begin{align*}
\text{if} & \quad \text{the SCOUR ceiling is exceeded in reactor 1} \\
& \quad \text{and} \\
& \quad \text{the SCOUR ceiling is not exceeded in reactor 2} \\
& \quad \text{and} \\
& \quad \text{the SCOUR ceiling is not exceeded in reactor 3} \\
& \quad \text{and} \\
& \quad \text{the SCOUR ceiling is not exceeded in reactor 4} \\
\text{then} & \quad \text{reduce flow rate into reactor 1} \\
& \quad \text{increase flow rate into reactor 2} \\
& \quad \text{increase flow rate into reactor 3}
\end{align*}
\]

The use of the flag enables us to shorten this rule to:

\[
\begin{align*}
\text{if} & \quad \text{the value of FLAG is 1} \\
\text{then} & \quad \text{reduce flow rate into reactor 1} \\
& \quad \text{increase flow rate into reactor 2} \\
& \quad \text{increase flow rate into reactor 3}
\end{align*}
\]

As previously mentioned, the value of SCOUR in each reactor is compared for each control period with the ceiling in each
reactor and a flag is set to a number between 0 and 15. There is a specific control action for each process state and these are synthesized into a system based on 16 different rules of operation.

Figure 3.19A shows the results of simulations for a base case in two dimensions. The SCOUR controller was not enabled during this simulation and thus all the influent flow was routed to reactor 1. Figure 3.19B shows the results with the controller active. The same influent flow rate and concentration signals were used for both simulations. Two things can be concluded from comparing these results:

- the variability and magnitude of SCOUR in the first three reactors was significantly decreased
- the value of SCOUR in the last reactor was not increased as a result of control

The controller is, therefore, achieving the goals defined earlier.
Figure 3.19  Results of SCOUR profile control
3.3.2.3 Simulation of Overall Strategy

In order to demonstrate the effects of changes in different variables on operation of the biological reactor, several conditions were simulated. A sinusoidal signal with a mean value of 450 cu. m./hour, an amplitude of 250 cu.m/hour, and a period of 24 hours was used for the influent flow rate signal for all simulations described in this section, unless otherwise noted. The volume of the biological reactors was 2500 cu.m total, making the average hydraulic residence time \( t = 5.55 \) hours.

First, a base case was defined to serve as a reference for comparison. Air flows to each biological reactor were the same throughout the simulation, the DO concentration was not controlled and all of the influent flow rate was always directed into the first reactor. The DO concentration in each reactor for one day of simulation is shown in Figure 3.20B.

It can be seen from this figure that the DO concentration in the first reactor decreases to low values during the afternoon hours when the plant is subject to increased loadings. Figure 3.20A contains the values of SCOUR in each reactor for the same simulation run. When compared with Figure 3.20B, it can be concluded that the microorganisms are subjected to low dissolved oxygen concentrations during the periods of highest biological activity.
A) Variation in SCOUR

B) Variation in DO

Figure 3.20 DO and SCOUR dynamics in a no-control case
The energy required by the compressors during 24 hours for this type of operation was 10,335 kWhrs.

With the other parameters remaining the same, the effects of DO control are examined in the second simulation. The DO set points were 2 mg/l in each reactor. DO concentrations were used as signals for manipulating the air flow rate to each reactor. Results of the DO control simulation for a 24 hour period are shown in Figure 3.21.

During all the simulations, the details of the control strategy for the compressors as described in chapter 2.6 were kept the same. The simulations show significant savings in energy for the controlled case vs the uncontrolled case since the DO is maintained at a substantially lower level. This is as expected since it is well proven (15) that DO control can reduce energy consumption in the activated sludge process.

The next set of simulations examine the effects of different set points for dissolved oxygen on energy consumption by the compressors. Table 3.4 shows the results of the simulations.

Energy utilization decreases with even a small difference in DO set points. Since the transfer of oxygen from the air into the water is proportional to the difference between the actual and the saturation concentration, it is advantageous to keep the DO concentration as low as possible subject to
Figure 3.21 Results of DO control
Table 3.4  
Effects of D.O. Set Points on Energy Utilized for Aeration

<table>
<thead>
<tr>
<th>Setpoints for D.O.</th>
<th>Reactor Number</th>
<th>Energy Used By Compressors in kWhrs</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>1</td>
<td>2</td>
</tr>
<tr>
<td>Run 1</td>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>Run 2</td>
<td>1.0</td>
<td>1.5</td>
</tr>
<tr>
<td>Run 3</td>
<td>2.0</td>
<td>2.0</td>
</tr>
<tr>
<td>Run 4</td>
<td>2.0</td>
<td>3.0</td>
</tr>
<tr>
<td>Run 5</td>
<td>2.0</td>
<td>3.0</td>
</tr>
</tbody>
</table>
the constraints of adverse effects on biological activity or inadequate mixing for maintaining the sludge in suspension in the reactor.

The effects of different MLSS concentrations on energy consumption were examined by simulating plant operation at different MLSS concentrations. Figure 3.22 shows the energy consumption obtained in these simulations. For all the simulations it was assumed that the dissolved oxygen control was in operation and maintained the DO concentration at 2mg/l in each reactor.

Aeration energy requirements were not significantly influenced by the SCOUR controller. During a 24 hour simulation period, the energy requirement for a case without SCOUR control was 8,982 kWhrs, compared to 9,050 kWhrs for a 24 hour period with SCOUR control, an increase of less than 2%. In both cases, the DO concentration was maintained at 2 mg/L in each reactor.

Air supplied to the biological reactor serves the dual purpose of providing enough oxygen for the biological reaction and enough mixing to keep the floc in suspension. As the hydraulic residence time in the biological reactors is increased, a point will be reached at which DO control becomes ineffective because the demand for air is controlled by the mixing requirements. Several simulations were performed to examine the influence of mixing requirements on dissolved oxygen control. The load to the plant was assumed
Figure 3.22 Effects of MLSS concentration on energy requirements
to be constant (the same mean value and amplitude of influent flow rate and concentration) and the hydraulic retention time of the biological reactor was increased.

Figure 3.23 shows the results of a simulation with the hydraulic retention time increased from 5.55 to 12 hours. The mixing requirements used in the simulations (25 cub.m.air/1000 cu.m.tank min) were adopted from the literature (101). Since the mixing requirements control aeration requirements, there is an increase in DO concentration in the last 3 portions of the reactor. This, of course, means more energy has to be utilized. In this simulation, energy consumption was 11,450 kWhrs.

Such a situation was at times observed at the Sagemont plant which normally operated well below its design load. The DO concentrations in passes 3 and 4 were sometimes quite high, but at the same time clear layers on top of the basins indicated inadequate mixing.
mean hydraulic retention time = 12 hours
DO setpoint = 2.0 mg/L in each reactor
mixing requirements = 25 cu. m air/1000 cu. m tank

Figure 3.23 Effects of mixing requirement limitations on DO control
3.4 Solids-Liquid Separator

The solids-liquid separator is not a likely candidate for the application of classical feedback control. Two variables which influence the separator and can be changed in a reasonably short time, and in a straightforward fashion, are the feed point (step feed) and the recycle flow rate. There is, however, controversy surrounding the use of the recycle flow rate as a manipulative variable (45). Although in this research the recycle rate was not manipulated during normal operating conditions, it should be further examined from the control standpoint. Severin et al. (102) have proposed a method for determining the minimum required underflow rate for given loading conditions in order to avoid blanket failure. Their method is based on solids flux theory and was tested on an 18 MGD industrial wastewater treatment plant. The author feels that the use of on-line predictions of the influent flow rate in conjunction with this method should be explored in future research.

3.4.1 Factors Affecting Separator Performance

The performance of the solids-liquid separator can be expressed by its ability to consistently produce an effluent containing low concentrations of suspended solids. As mentioned in section 3.4, there are two groups of mechanisms affecting clarification and thickening - physical (hydraulic) and biological. The biological aspect is
reflected in characteristics of the sludge. It is proposed that the objective of the control strategy should be to establish and maintain the most favorable conditions, from both the physical and biological aspects, for operation of the solids-liquid separator. The role of the control system will not be to react in a feedback manner to perturbations in the output signal from the separator (such as effluent suspended solids), but rather to continuously provide operating conditions in the biological reactor most likely to have a favorable effect on the operation of the separator. These conditions are defined as:

- Conditions most likely to lead to the production of solids with good settling properties (biological aspects)

- Conditions in which the least amount of turbulence is created in the separator (physical aspects)

The operation of the biological reactor will influence the settling properties, and therefore the control strategy for the biological reactor has as an objective the production of solids with good settling properties. This control strategy is applied through the SCOUR controller and is described in Chapter 3.3.

3.4.2 Variation of Flow Rate

As noted earlier, it is not possible to significantly reduce
the variation in the influent flow rate by control of the pump station, since the volume of the wet well is not large enough to provide the needed dampening capacity. Turbulence caused by variations in flow rate from the pump station should not have any adverse effect on the operation of the biological reactor since this is a turbulent environment, anyway. Reducing the variation of flow rate into the separator, however, should improve its performance. Variation of the level in the biological reactor could be used to dampen the variation in the flow rate to the separator since the volume of the biological reactor is significantly larger than the wet well volume.

Poduska (103) reports experiences with an industrial wastewater treatment plant (average flow 2800 cu. m/hour) in which the clarifiers were constructed at a higher elevation than the biological reactors. Because of the difference in elevation the clarifier influent had to be pumped from the biological reactor. This created an opportunity to control the flow between the biological reactor and the separator. Poduska allowed the level in the biological reactor to vary 6 inches, thus providing 0.75 MG capacity to dampen the variations in the load to the separator. This method of flow equalization nearly eliminated loading variations on the clarifiers. As benefits of this control strategy Poduska mentions fewer operating difficulties in controlling sludge blanket levels and a reduction in the effluent suspended solids.
The controller proposed is a PI controller which uses the level in the biological reactor as a process variable. The level in the reactor was allowed to vary 0.3 meters. The flow out of the biological reactor was proportional to the level in the basin. Figure 3.24B shows the outlet flow, which may be compared with the influent flow signal shown in Figure 3.24A. Most of the high-frequency noise is dampened.

The goal of the flow controller is not to maintain tight level control, but rather to provide smoothing of the outlet flow rate. This problem is referred to as the averaging level control problem, and has been extensively investigated in the chemical engineering process control literature. Solutions have been proposed by Luyben and co-workers (104-107), and Shunta and Fehervari (108) and Cutler (109). McDonald and McAvoy (77,110) discuss an implementation of predictive control, and compare the results with those obtained by other types of controller.
Figure 3.24  Results of flow control between the biological reactor and the separator
4. SYSTEM CONTROL

4.1 Approach to System Control

Up to this point, control of the process during routine operation has been examined separately from responses to large disturbances. An integrated control strategy would connect all the control loops into a comprehensive and coordinated system that manipulates the operation of all the units during any process conditions. The task of the integrated control strategy is to assess the state of the process on-line, and determine whether significant disturbances have occurred. In the case of upset conditions, it calls on appropriate special control strategies to reduce the effects of the upsets on plant performance and return the process to normal conditions.

An integrated control strategy acts as an "umbrella" for system operation. For example, algorithms can be developed which are capable of detecting faulty measurements, unusually high flow rates, a rising sludge blanket, or other significant conditions. This information is used along with other information collected from the sensors to determine the general state of the process. If this overall state of the process is not an upset condition, all the control loops operate normally. In the case of an upset condition, the appropriate action is taken through the use of special procedures in order to return the system to its normal
operating range.

It should be noted that an important part of the integrated control strategy was not considered in this work - this being the interface between the operator and the process. At least the first generation of integrated control systems for the activated sludge process will have to include a human operator in some of the control loops. The problem of an adequate operator-process interface is a complex one, and there is significant room for improvement in this area. It is suggested that this aspect should be examined in future studies of the control of activated sludge and other biological processes for wastewater treatment.
4.2 Interactions Between Process Units

Interactions between the process units are identified and discussed in the following paragraphs. The list of interactions is by no means complete. Interactions should be considered in both operation and design of activated sludge treatment plants.

Interactions have not been extensively studied in the past although experienced operating engineers are aware of their importance. It is necessary to have a comprehensive model of the process in order to investigate the interactions on a quantitative basis. Developing a comprehensive model is a complex task since the mathematical models for individual process units are still under development and model validation is still incomplete.

Interactions exist between the process units, between plant design and operation and between control loops. The importance of these can hardly be overstated. A specific example of design-operation interactions is the interaction between the design value for the hydraulic retention time of the biological reactor and DO control. Simulations performed in this study confirm that aeration in plants with long retention times is limited with respect to mixing in the later stages of the reactor. DO concentrations must therefore be higher than those required for biological activity in the last stages of the biological reactor.
4.2.1 Solids-Liquid Separator and Pump Station

During the development of control strategies for the solids-liquid separator it was stated that there were two basic groups of mechanisms affecting the thickening and clarification functions of the separator.

The first group of mechanisms are physical in nature. Increases in the effluent suspended solids concentration have been observed to be correlated with rapid increases in influent flow rate to the separator. The research of Olsson and Chapman (67) is an example where physical mechanisms have been found to be significant and perhaps the dominating factor in performance. In order to avoid this problem, it is necessary to reduce the rates of change of flow rate into the separator, especially sharp increases.

It has been shown earlier that, due to the small size of the wet well, it is not possible to significantly dampen variations in flow rate. However, volume for equalizing can be found either in the sewer network, or, as proposed here, in the biological reactors. From the control standpoint, the problem is identical. The same basic control approach can be used whether the volume in the sewers or the volume in the biological reactors is used to dampen the variation in the influent flow rate.

With flow rate controlled between the biological reactor and the separator, operation of the lift station does not affect the clarification process. The identification of
this interaction and an understanding of its implications led to an alternate operational strategy which has significant potential for improving performance at low cost.

If control of flow rate between the biological reactor and the separator through the use of a variable volume reactor is not possible, the operation of pumps can have an effect on clarification. Although On/Off control of pumps was shown to be more energy efficient than the use of the variable speed or automatic throttling pumps, deterioration of effluent quality might result from hydraulic shocks that occur when a large pump is turned on.
4.2.2 Solids-Liquid Separator and Biological Reactor

The second group of mechanisms affecting separator operation are biological in nature. While the physical environment in the separator is important, the settling properties of the particles are at least equally as important. Both thickening and clarification are strongly affected by the settling properties of the activated sludge. Several investigators (96-99) have explored the effects of the "history" of the sludge on settling properties, and the conditions under which biological growth occurs are recognized to have a pronounced effect on sludge characteristics.

It is assumed in this research that the reason for the deterioration of settling properties of the activated sludge is excessive growth of filamentous organisms due to an oxygen deficiency although it is recognized that this is only one of many factors which influence settling characteristics. The settling properties are therefore controlled by creating the most favorable conditions for development of a sludge with good settling characteristics. The fact that the operation of the biological reactor is controlled in order to improve the operation of the separator shows the strong interaction between the two units.

The biological reactor and the separator are linked by the recycle of the activated sludge. This link is well
recognized as an important interaction but its effect was not explored in this study except for use during upset conditions. The recycle flow rate was maintained at a constant value during normal operation.
4.3 Interactions Between Controllers

The control loops addressed in this research include:

- **MLSS controller.** The waste sludge flow rate is controlled on the basis of MLSS concentration in the last reactor, or total mass of solids in the system.

- **Dissolved oxygen controller.** The air flow rate to each basin is manipulated to maintain DO concentrations at a set point in each reactor.

- **SCOUR controller.** A near constant value of SCOUR is maintained in space and time through control of the contacting pattern between the raw wastewater and the activated sludge.

- **Flow controller.** The flow rate between the biological reactor and the solids-liquid separator is controlled. Variation in biological reactor volume is allowed within upper and lower limits.

- **Compressor control.** The pressure in the main header is maintained at a set point.

- **Pump station control.** The level in the wet well is used as a signal, and pumps are operated to keep this level within upper and lower limits.

All of the control loops mentioned above act concurrently, and in some cases the output of one controller affects the
input to another controller. The interactions between loops have to be considered in the overall control strategy design, in order to avoid possible detrimental effects on performance. Several examples of such interactions will be presented in this chapter.

In order to improve clarification in the solids-liquid separator, a control strategy has been proposed to control the flow between the biological reactor and the separator. The level in the biological reactor is allowed to vary thus providing volume for dampening.

Although it is the performance of the separator that is being influenced the control action is performed in the biological reactor, with the purpose of decreasing the effects of turbulence created by the pump station. The allowed variation in the liquid level in the biological reactor is only 0.3 meters, but it can affect the aeration system. The head against which the compressors operate changes with a difference in depth, therefore affecting compressor control. The compressors are controlled to maintain a constant pressure in the main air header. The set point for the pressure in the main air header may be decreased if the liquid level drops in the reactor, since there is less static pressure to overcome.

Interactions between the DO controller and the SCOUR controller are shown in Figure 4.1. The upper plot shows the results of DO control when SCOUR control was not active
SCOUR controller not active

SCOUR controller active

Figure 4.1 Interaction between SCOUR and DO control loops
and the bottom plot show the results of DO control when the SCOUR controller was active. The difference is not significant.

There will definitively be an interaction between the MLSS controller and the solids processing train. The condition and concentration of the waste activated sludge, for example, affect the solids processing train, so both operation of the biological reactor and the solids processing train need to be considered by the waste sludge controller. This interaction should be examined when the system is expanded to include the solids processing train.
4.4 Interactions Between Design and Operation

Treatment plant design obviously has a strong influence on the choice and performance of a control strategy. The design of most plants is based on steady state criteria, or at most maxima and minima, whereas during plant operation performance is normally dynamic. Also, there may be inadequate communication between the design and control engineer during the design phase. If a control system is designed for an existing plant, the constraints imposed on the control engineer by the characteristics (size of the units, flexibility etc.) of the plant may mean the difference between success or failure of the control system. Lack of incorporation of step feed capability, for example, limits possible control strategies for protection against failure of the settler or control of dissolved oxygen.

There are two main concerns in the application of control to a treatment plant:

- How dynamic is the operation of the plant (without control)?

- How flexible is the plant? Flexibility means that process variables and plant configuration can be manipulated.

The treatment plant at Sagemont (Appendix 8.1) may be used as an example to illustrate these concerns. This plant was significantly underloaded so that hardly any dynamics could
be observed in normal operation. The plant was large enough to dampen most variations in load and thus little, or any, improvement could be made in effluent quality which was consistently better than the permit requirements. With two compressors running, several inches of an almost clear zone in the last 2 of the 4 reactors indicated that the air supplied was barely adequate for mixing. Still, the DO concentrations in these reactors were high most of the time because of the low oxygen demand. Efforts to reduce energy utilization were hindered by the second concern - plant flexibility. Air supply distribution could not be manipulated between the reactors, therefore restricting the amount of control that could be applied. This combination of the lack of dynamic performance and limited flexibility created difficulties in experimental exploration of possible control strategies. Partial step feed capability was added to the plant in order to increase flexibility and was shown to be a valuable tool in operation.

The above limitations resulted in a major portion of the research reported herein being conducted by computer simulation using a dynamic model to represent the plant. In order to investigate various control strategies, it was assumed that the plant was fully instrumented and that all final control elements could be remotely controlled. A short summary of important features which would give plants more flexibility and allow easier implementation of advanced control is presented in this chapter.
Recycle flow rate was not investigated as a control variable but was instead maintained at a constant value in all the simulations of normal plant operation. Recycle flow rate was only changed during special procedures in order to simulate a response to significant disturbances.
4.4.1 Step Feed

The capability of controlling the distribution of influent flow between biological reactors is essential for the control strategies proposed herein. As mentioned earlier, the benefits of step feed were first established by Torpey(86) and have since been studied by Sorensen(87), among others. In this study, step feed is used to control SCOUR in the biological reactors and is also used as a part of a special procedure for preventing process failure.

4.4.2 Control of Air Flow Distribution

Due to differences in oxygen uptake rate the demand for oxygen varies from stage to stage in the biological reactor. If all the wastewater is introduced into the first stage, changes in SCOUR are significant. The uptake rate is largest in the first stage, and decreases towards the last stage as shown in Figure 3.16. This fact has been recognized for some time, and many plants have aeration systems designed to "taper" the aeration along the length of the basin.

The aeration system is the largest single "user" of energy in an activated sludge plant, and dissolved oxygen control can be used to minimize the energy required to supply air to the biological reactor. The most effective DO control can only be accomplished if both air flow rate and distribution can be controlled. It is therefore important that
provisions be made for this during the design stage.

4.4.3 Control of Flow Between the Biological Reactor and Solids-Liquid Separator

Variation in the biological reactor volume can be used to reduce variations in flow rate to the solids-liquid separator, and thus avoid possible deterioration of effluent quality. This control strategy provides better dampening of influent flow rate than the use of variable speed pumps. In treatment plants with several solids-liquid separators in parallel, placing control elements in the influent to the solids-liquid separators would reduce variations in flow rate as well as provide a more equal distribution of flow to each unit.

4.4.4 Instrumentation

It is important to adopt control strategies that use measurements from reliable and easily maintainable instruments. The control strategies proposed herein are based on instruments such as dissolved oxygen meters, flow meters, and level indicators, which are relatively reliable and require only limited maintenance. Measurements were assumed to be noise-free for the purpose of this research. The handling of erroneous measurement readings is discussed in the chapter on special procedures.
4.4.5 Computing Hardware and Software

In order to implement the proposed control strategy, the plant would have to be equipped with a process control computer and/or programmable controller. These control strategies require on-line information from the sensors and the processing power and memory of a computer.

4.4.6 Operator-Process Interface

Modern technology can provide wastewater treatment plant operators with many useful tools. Data acquisition systems and computer generated graphics have a great potential for making it easier for the operator to monitor and control the process. The ability of the computer to store and retrieve data on request can provide the operator not only with the present values for the variables, but, by the use of graphics and charts, with an effective tool to detect trends and patterns from the recorded performance.

The flow of information goes two ways. While the operator is able to monitor the operation through the information supplied to him by the monitoring and control system, he is also able to provide information that is difficult to obtain via sensors. An "user friendly" interface should be able to obtain information from the operator (odor intensity, for example), and pass it on to the control system.
4.5 Special Procedures

The control strategies described in this research are designed to control the plant during "normal" operating conditions, which are defined as conditions in which no event is detected that might cause process failure. If a disturbance is detected, special procedures are activated in order to avoid a significant process upset or failure. Several procedures are listed and described in this section.

4.5.1 Toxic Input

One example of a disturbance which can cause process failure is the presence of toxic substances in the influent. This condition occurred on two occasions at Sagemont when a valve ruptured on the bleach storage tanks, causing the bleach to drain into the wet well. If the bleach is pumped into the biological reactor, it can result in death of a significant fraction of the microorganisms and thus system failure. If the plant had been equipped with instruments to measure the flow of bleach out of the tank and the level of bleach in the tank, an accident like this could have been quickly detected. A simple mass balance would have indicated that the drop in level did not correspond to the flow rate measured at the outlet. The computer could have raised an alarm and/or initiated corrective action to alleviate the problem.
In a fully instrumented plant with a data acquisition system and a process control computer it is possible to design programs to detect unusual changes in process variables that may indicate the presence of a toxic substance. For example in the event of a toxic input to the plant, a rise in DO concentration and a drop in SCOUR would be observed.

Once a toxic input is detected, the plant may be controlled in accordance with rules set up for the specific situations. If off-line storage is available, influent can be temporarily routed to this. It can be later diluted and slowly introduced into the plant. If such storage is not available, the recycle flow rate can be temporarily stopped and the influent flow feed point switched to the last reactor. This control action would minimize the exposure of the microorganisms in the biological reactor to the toxic substance, thus preventing a long lasting detrimental effect on the process, but could result in temporary violation of the permit limits for the effluent. Regardless of what control action is determined to be most suitable, it is important that it be executed in a timely manner.

There are many factors to be taken into consideration concerning the appropriate response to such a disturbance. The magnitude of the disturbance and the nature and the condition of the receiving body of water, are only a few of the items which must be taken into consideration in determining the optimal control action.
4.5.2 Hydraulic Shock

During storm events, there can be substantial increases in the influent flow rate, and the pattern of the influent flow rate vs. time may change drastically from the dry weather pattern. Figure 4.2 shows typical dry and wet weather flow rate periods at the Sagemont plant.

Since the influent flow was well below design capacity during the study at Sagemont, it was possible to take a portion of the plant out of service thus reducing the energy required for aeration. This reduction was possible because the aeration was mixing, not reaction limited. However, to avoid excessive discharge of solids, provision was made to automatically bring an out-of-service settler back on line when storm flows exceeded 15 MGD. This control action was responsible for the sudden decrease in flowrate that occurred at 6 AM on the storm day shown in Figure 4.2b.

This illustrates that it is possible to detect a significant departure from the usual flow pattern by using the ability of the process control computer to store the history of the influent flow rate. Data from rain gauges would be usefull in predicting increases in flow rate ahead of time, but even a simple interactive program that would allow the operator to pass information about the weather to the data
Figure 4.2 Dry and storm day flow rate at Sagemont plant
acquisition system would make it much easier for the control system to deal with the problem and for the control programs to "understand" the changes in the process.

4.5.3 Failure Due to Rising Blanket

In a normally loaded plant the sludge blanket rises and falls due to diurnal variations in loading. Sometimes, however, it may reach the point at which there is a gross discharge of solids over the weirs. It is necessary to detect this condition and exert control to prevent this from occurring. Simulation of such control is presented in this section.

The mathematical model for the solids-liquid separator predicts the solids concentration profile in the separator, and was used to test the control strategy. The solids-liquid separator is divided vertically into ten layers, and the concentration of solids in each layer is calculated. In order to obtain the same amount of information about the solids profile in an actual separator, it would be necessary to have either a moving head solids sensor, or several fixed solids sensors at various levels. It is assumed in this study that only a single solids sensor is available. The position of the blanket detector is in the third layer (the first layer is at the top of the separator.).

If the sludge blanket reaches the detector this is treated as a disturbance, an unusual condition that arises due to an
overload of solids to the separator. During routine operation, when the sludge blanket is not high, no control action is performed. If the sludge blanket reaches the third layer, the recycle sludge flow rate is increased and the contacting pattern is changed. As a result of this control action, the solids concentration rises in the first reactor and declines in reactors 2 through 4. The load to the settler decreases, and solids are stored in the first reactor. When the sludge blanket drops, the contacting pattern is once again determined by the SCOUR controller. Figure 4.3 and Figure 4.4 show the results of a simulation during which influent flow rate into the plant was sharply increased for a duration of 4 hours. Figure 4.3 shows the changes with time of the concentration profile in the separator. It can be seen that the increased load causes a rise in the sludge blanket, but because of the scale on the vertical axis it is not apparent from Figure 4.3 what was the effect on the effluent suspended solids concentration. Figure 4.4. shows the concentration of solids in the top layer of the clarifier during the same simulation, and since the concentration within each layer is assumed to be homogenous, this is also the effluent suspended solids concentration. Figure 4.5 shows the concentration profile in the separator obtained from a simulation during which a special procedure was activated. The control was activated when the solids concentration in the third layer exceeded the set point.
Figure 4.3 Solids profile dynamics in the separator
Rising sludge blanket due to sudden increase in loading
No control case
Figure 4.4 Solids concentration in the upper layer of the separator. Rising sludge blanket due to sudden increase in hydraulic loading.
Figure 4.5 Solids profile dynamics in the separator
Rising sludge blanket due to sudden increase in loading
Special procedure activated
Recycle flow rate was increased by 100%, and the influent flow rate was divided between the first reactor (which received 10% of the influent flow rate) and the second reactor (which received the remaining 90%). This resulted in solids being stored in the first reactor instead of accumulating in the separator. It can be seen that the sludge blanket did not continue to rise and there was no increase in the solids concentration in the top layer of the separator.

4.5.4 Erroneous Sensor Readings

The control system "sees" the process through measurements obtained from the sensors and its perception of the process and ability to determine its state are dependent on the accuracy of the sensor readings. It is therefore important that the measurements have the appropriate accuracy.

Three different problems will be briefly discussed here - the effects of instrument calibration, sudden instrument failure, and gradual deterioration in accuracy (drift).

The first problem arises during the calibration of instruments. For example, the DO meter reading will change drastically during calibration, as shown in Figure 4.6. It is important to recognize these changes as being brought about by the calibration procedure rather than by some significant change in the process state. One possibility is to set a flag in the data base before calibrating and
Figure 4.6 Example of DO signal during calibration

January 23, 1983
Aeration Basin 3
clear the flag after the calibration is accomplished. The computer will thus know if calibration is being performed when it encounters sudden unexpected changes in the signal and disregard the values in the signal during calibration. Automatic detection by a computer program using pattern recognition techniques is also feasible.

The second problem is sensor malfunction. During the Sagemont project it was observed that much of the time instrument failures frequently resulted in no signal from the sensor (0 mA, rather than a number between 4 and 20 mA) or in a "stuck" signal.

The signal was called "stuck" when a sensor was sending a fixed value, even though the variable was changing. Since the effluent flow meter exhibited this behaviour quite frequently (Figure 4.7), a program was developed to determine and report the problem. The program used a subroutine that compared the difference between the successive one-minute readings for the last 15 minutes to determine if the signal was likely to be valid. The program subtracted the successive measurements in time and counted the number of times the difference was below the established tolerance value. If the number of occurrences exceeded a certain value, it was likely that there was a problem with the sensor. This program was tested on-line and performed without any problems.
Figure 4.7 Effluent flow meter failure
Another example of such a failure was experienced in the pump station. In this case, the wet well level sensor failed and indicated the wet well level to be more than 7 feet while actually the level had dropped below the low alarm point. Since the entire control strategy in the pump station is based on the wet well level measurement, it is very important to detect sensor failure and prevent the control strategy from responding to an erroneous signal. A computer makes it possible to deal with such a problem in a rather simple and straightforward fashion. In addition to using the same approach as the one used on the effluent flow meter, it is also possible to utilize the information obtained from the other sensors to check the validity of the level signal. Discrete pressure switches are used to determine the levels at which the pumps turn on and off and can be used to verify the continuous level signal. Flow and pressure measurements at the outlet of the pump can also be used to detect this problem.

A third type of error occurs over a long period of time and is commonly referred to as "drift" in the sensor. If an increasing bias is detected by comparing the measurement with the other measurements or its long term history, a request for calibration may be issued to the operator.
5. CONCLUSIONS

From the results of this study it may be concluded that:

1) It would be desirable to define and estimate the state of the activated sludge process on-line for the purpose of process control. Many variables routinely used in control of the activated sludge process, such as the DO concentration, do not contain sufficient information to establish the process state and should be treated only as constraints. Process control computers may be used for processing the information obtained from a combination of sensors and to estimate parameters, such as SCOUR, that contain more meaningful information about the process. A combination of SCOUR and DO concentration is proposed for estimating the process state.

2) The state of the activated sludge process is difficult to define, thus making this process difficult to control using conventional control systems. Rule based controllers are well suited for systems where definition of process state is complex and where much of the information about the process is qualitative. Rule based controllers show good potential for control of the activated sludge process. Simulations of the rule based control system proposed in this study indicate that it has significant potential for
controlling SCOUR in both space and time.

3) Deterministic and stochastic models can be successfully combined in a comprehensive model for the activated sludge process. An optimally balanced model should take advantage of both modern statistical methods and available a priori knowledge. Time series methods were successfully used in this study to predict the plant influent flow rate one hour in advance. The adverse effects caused by the nonstationary nature of raw data were eliminated by pretreatment of data. Flow data collected at the Sagemont plant were used to test the flow prediction model. The prediction error was found to be within 10 to 15% of average flow rate on a dry weather day.

4) The thickening model proposed in this study successfully predicts the solids profile in the solids-liquid separator based on data published in the literature. This result suggests that the model can be used to predict the effect of control actions on the solids-liquid separator. The simplicity of the model makes it suitable for use within computer-based real-time control systems.

5) A strong stochastic component was observed in the effluent suspended solids concentration at the low values observed in the Sagemont study. The use of strictly deterministic models for predicting low values
(less than 10 mg/L) of suspended solids is therefore questionable.

6) The wet well volume in pump stations may be insufficient to provide the necessary capacitance for dampening of influent flow rate and thus decreasing variations in flow rate to the solids-liquid separator. The use of a variable volume biological reactor has the potential for significantly reducing these variations in flow rate. This control strategy would also permit improved control of distribution of the flow rate between the individual solids-liquid separators.

7) For low-friction-head, high-static-head pump stations, On/Off pumping requires less energy than variable speed pumping or automatic throttling of the discharge. On a typical dry day, 904 kWhrs and 201 kWhrs were used at the Sagemont plant for the variable speed and On/Off pumping, respectively.

8) Strong interactions exists between design and operation of the activated sludge process. Possible improvements in operation of an existing plant through implementation of process control are highly dependent on the plant design. For example, the SCOUR controller proposed in this study utilizes step feed to maintain the desired SCOUR profile. It is therefore important that plants be designed to include step feed capability. Another example is the ability to control
the air flow rate to each aeration basin. This capability allows the control engineer to maintain the DO at the desired level in each basin and meet the aeration demand in an optimal manner.

9) The interactions between control loops have to be considered in the overall control strategy design in order to avoid possible detrimental effects on performance. For example, there will definitively be an interaction between the MLSS controller and the solids processing train. The condition and concentration of the waste activated sludge affect the solids processing train, so both operation of the biological reactor and the solids processing train need to be considered by the waste sludge controller.

10) The objective function for the control system depends on the state of the process, and therefore the control strategy has to be a function of the system state. Two different objective functions were identified and analyzed in this study. The first objective function is to maintain adequate wastewater treatment at minimal energy cost, and the second objective function is to avoid process failure in the event of significant disturbances to the process. The control strategy designed to minimize the first objective function is proposed for use during normal operation conditions. Special procedures are invoked
if a significant disturbance to the process is detected and the objective is to avoid process failure.
6. RECOMMENDATIONS FOR FUTURE RESEARCH

As is the case for many research projects, more questions were created than were answered. This clearly indicates that further research is needed in modeling and control of the activated sludge process. The significant portions of both the model and control strategies need to be verified with experimental data. Several examples where improvement is needed will be presented:

1) The model used in this study predicts effluent suspended solids concentration during upset conditions but does not include prediction of effluent suspended solids during normal operating conditions. The data from the Sagemont plant indicates the presence of a strong stochastic component for this variable. For this reason, the time series method used in this study for influent flow prediction should be suited for effluent solids prediction. It is recommended that this approach be evaluated.

2) The micro conditions in the biological reactor were defined in this study to be a function of only the DO concentration and SCOUR. Additional factors, such as the nutritional balance of the influent substrate, should be quantitatively investigated and included in future control strategies.

3) Future investigations should expand the size of the
system studied by including the solids processing train, sludge disposal site, sewer network and the receiving stream so that interactions between these components may be more clearly defined and studied. This may be done by blending models for these components with the model presented herein.

4) Additional research should be conducted on establishing reliable indicators of process state from on-line information. Additional methods for estimating important process variables such as SCOUR, the oxygen transfer coefficient and sludge settling characteristics should be explored.

5) The development of fuzzy set membership functions for important process variables, such as SCOUR, DO, and MLSS would be an important step in the design of control systems using qualitative information. Fuzzy set theory provides mathematically rigorous methods for handling such qualitative information. It is recommended that application of these tools be studied for control of the activated sludge process.

6) A large body of empirical knowledge is available from experienced operating engineers. This information would be valuable for development of rule based control systems for the activated sludge process. It is recommended that this knowledge be collected and formalized so that it may be used for expert control.
7.0 REFERENCES


46. Louis Allis, Inc., "Maintenance Manual for Louis Allis Lancer 44XLP Simplex Drive System for Worthington 6-860640 125 HP."


8.0 APPENDICES

8.1 Description of Sagemont Treatment Plant

Data collection and experimentation were carried out at one of the City of Houston's wastewater treatment plants. For research purposes, the plant was instrumented with on-line sensors and controllers. The data acquisition / data base system consisted of a minicomputer (DEC PDP 11/23) with mass storage devices and three industrial programmable controllers (TI PM550) for interfacing the field devices with the computer.

The Sagemont wastewater treatment plant is located in the southeast part of Houston, and receives largely domestic wastewater from the surrounding residential area. Figure 8.1 shows a plan view of the plant. It was constructed in the 1960's and expanded to its current capacity in 1975. The plant has been in continuous service since the expansion in 1975.

The design dry weather capacity of the plant is 5 million gallons per day (MGD) with a maximum hydraulic capacity of 25 MGD. However, with an average daily dry weather influent sewage flow rate of around 2 MGD, the plant is considerably under-loaded and thus the effluent quality was very good. Energy consumption was high due to excess air being supplied to the activated
Figure 8.1 Plan view of Sagemont treatment plant
sludge process. In order to achieve a plant loading closer to the design value, approximately half of the plant was taken out of service. This modified flow scheme is shown in Figure 8.2.

The influent to the plant is pumped through three pump stations. One is located on site while the other two are located off site. The pumped wastewater enters the bar screen pit, where coarse floating materials are removed. It is then mixed with the return activated sludge immediately after entering the first aeration basin. The mixture of return activated sludge and wastewater, i.e. the mixed liquor, flows sequentially through aeration basins 1, 2 and 3, and the aeration channel as shown in Figure 8.2. The typical hydraulic residence time in the aeration basins is 10 to 12 hours (based upon influent flow rate).

The mixed liquor is subsequently introduced to clarifier No. 2 which was also instrumented for research purposes. In the clarifier, solids are separated and pumped to the return sludge channel while the clarified liquid flows to the chlorination basin where it is disinfected using a solution of sodium hypochlorite.

Excess sludge is withdrawn from the second aeration basin for about 5 hours a day at a flow rate of approximately 500 gallons per minute (GPM). The excess
Figure 8.2 Configuration of the aeration basins for this study
sludge is pumped to aerobic sludge digesters for subsequent processing. The digested sludge is dewatered by a mobile filter press and hauled to a landfill site for final disposal.

Table 8.1 shows the dimensions of each unit and average operating conditions.

Table 8.2 shows the average influent flow rate and effluent quality for 1981 as reported by the City of Houston. Table 8.3 shows the effluent quality requirements (permit limits) at the plant.

The plant was modified to include feedforward of influent to aeration basin 3 for research purposes (Figure 8.1). A more detailed description of the plant can be found in a series of reports by Andrews (54-63).
### Table 8.1
Dimensions and Volumes of Aeration Basins

<table>
<thead>
<tr>
<th>Basin</th>
<th>Dimensions (m)</th>
<th>Volume* (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Aeration basin 1</td>
<td>9 x 37.4 x 5</td>
<td>1520</td>
</tr>
<tr>
<td>Aeration basin 2</td>
<td>9 x 33.5 x 5</td>
<td>1363</td>
</tr>
<tr>
<td>Aeration basin 3</td>
<td>6 x 29 x 5</td>
<td>795</td>
</tr>
<tr>
<td>Aeration Channel</td>
<td>6 x 68 x 3</td>
<td>1022**</td>
</tr>
</tbody>
</table>

*Based on a typical freeboard of .6 m.

**Actual liquid volume of the aeration channel is estimated to be roughly 1/2 of the calculated volume due to grit deposits in the channel.

### Table 8.2
Yearly Average Flow Rate and Effluent Quality (1981)

<table>
<thead>
<tr>
<th>Flow Rate (m³/hour)</th>
<th>Suspended Solids (mg/L)</th>
<th>BOD5 (mg/L)</th>
<th>NH3-N (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>360</td>
<td>6</td>
<td>3</td>
<td>0.2</td>
</tr>
</tbody>
</table>
Table 8.3
Permit Requirements for Sagemont Treatment Plant

<table>
<thead>
<tr>
<th></th>
<th>Maximum 7-day average (mg/L)</th>
<th>Maximum 30-day average (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Suspended Solids</td>
<td>20</td>
<td>12</td>
</tr>
<tr>
<td>BOD5</td>
<td>10</td>
<td>5</td>
</tr>
<tr>
<td>NH3-N</td>
<td>10</td>
<td>2</td>
</tr>
</tbody>
</table>
8.1.1 On-Line Instruments

About eighty analog measuring devices and fifty discrete monitoring devices were installed in the plant. The locations of some of the installed instruments are shown in Figure 8.3 and 8.4. In addition to the instruments indicated in these figures, both the compressors and the pump station were fully instrumented.

The field instruments were connected to one of the three Texas Instruments PM-550 programmable controllers, which served as interfaces to the main computer. The computer, a DEC PDP 11/23, was located in the Computer Monitoring and Control Center was interfaced to the programmable (PC-s) controllers or direct connection depending on the location of the PC.
1. DO meter
2. Suspended Solids Meter
3. Temperature
4. Flow Rate
5. Turbidity
6. Sludge Blanket Level
7. Manually Operated Valve (open)

Figure 8.3 Instrumentation for this study
8.1.2 Computer Monitoring and Control System

Figure 8.5 depicts the data acquisition and control software used at Sagemont. Signals transmitted to the main computer once every 6 seconds were filtered and/or averaged every minute, converted as appropriate and stored in the short history data file residing on one of the 10.4 Mega Bytes (MB) hard disks. This short history data file was formed as a queue and could hold the last 48 hours of data. The short history data file contained both engineering values and the status of discrete signals. The past 24 hours of data were transferred daily to permanent storage magnetic media.

The main data base and related programs were developed so that on-line control could be accomplished in addition to the monitoring and recording of system conditions. Since the research was carried out at an operating full-scale plant, limitations were imposed in the choice of manipulatable variables and their ranges.
Figure 8.5 Data acquisition and control software
8.2 Program Listings

8.2.1 Activated Sludge Model

This program was developed for simulation of the operation of an activated sludge process during dynamic conditions.

Aeration basin is approximated with four completely mixed reactors in series.
The units are grams, meters and hours.
Concentrations are expressed in oxygen equivalents.

Subroutines called by this program are:

INI - Initializes all the variables
GETTSS - Calculates MLSS in each reactor
WASTE - Determines waste flow rate
AIRSPL - Estimates Kla from airflows
AERSYS - Calculates power utilization by blowers
THKNR - Calculates underflow concentration from SLS
PARSUB - Supplies parameters and constants
INFL - Supplies influent flow and composition
FLOPRE - Obtains flow prediction
FLOW - Calculates influent flow pattern
DVERK - Diff. Eq. Solver from IMSL (single pr.)
SHRHIS - Creates very short history file
INTEGR - Obtains rough integrated values for PI
OTP - Outputs results to appropriate files

Data is mostly passed through common areas in memory.
X is the independent variable - time

```
INTEGER N,IND,IBR,NW,I,K,MAINFL,IBLO(4)
REAL Y(44),C(24),W(44,9),XI,TOL,XEND,YPR(44),Y1(44)
REAL OUR(44),CONS(16),SCOUR(44),DOX(44),AIRF(4)
REAL KSP,KFS,KD,KDNS,KDNB,KN,MHNS,RES(44),TSS(4)
REAL PAR(44),INTEG(16),DER(16),DOHAT(4),CEIL(4)
REAL FOR(44),FML(4),V(4),SHORT(52,3),KLA(4),FOLD(4)
REAL OLDPRF(30),HPOWER(4),CSET(10),SETP(4)
```

COMMON /CONTRL/DOK,SETP,SCOUR,AIRF,INTEG,DER
COMMON /FLOWS/,F01,F02,F03,F04,FWA
COMMON /TURN/,FR,XR
COMMON /OXUPT/,CONS
COMMON /OLDPRO/,OLDPRF
COMMON /PARAM/,RDT,FSHAT,RSD,RH,KSP,RXA,Y1,KFS,KD,
1 OHNS,KDNS,OHNB,KDNB,KN,Y2,FRACT,YNS,YNB,CNIT,CAM,Y1I
COMMON /INFLU/ F0,S0,SNH30
COMMON /YVECTR/ RES
COMMON /WASTFL/ TSS,FW
COMMON /VOLFLO/ FOR,FM1,V
COMMON /SHRTH/ SHORT
COMMON /AIR/ KLA
COMMON /PREDIC/ FHAT,DOSHAT
COMMON /PATERN/ FOLD,CEIL,MAINFL
COMMON /BLOPOW/ TDEOND,HPOWER,IBLO
COMMON /XPROFL/ CSET
COMMON /TURBID/ TUR

EXTERNAL FCN1

UNITIES ARE GRAMS, METERS, HOURS
CONCENTRATIONS ARE IN G OE/ CUBIC METER (MG OE/L)

SET UP PARAMETERS FOR DVERK

N = 44
NW = N

X = 0.0
TOL = 0.025
IND = 1

CALL INIT (N,Y)

GET INITIAL CONDITIONS

DO 65 K=1,44
RES(K)=Y(K)

65

CALL GETTSS(X)

CALL WASTE(X)

CALL AIRSPL (X)

CALL AERSYS (X)

GET KINETIC PARAMETERS

CALL PARSUB(X)

SPWR = 0.0

OPERATION IS SIMULATED FOR THE DURATION
OF LDAYS DAYS AND THE LAST DAY IS WRITTEN
OUT INTO THE APPROPRIATE DATA FILES

LDAYS = 7
LEND = LDAYS*96 + 12
DO 22 L=1,LEND

15 MINUTES (1/4 HOURS) INCREMENTS

XEND = FLOAT (L) /4.0
CALL INFL(X) GET INFLUENT FLOW AND CONCENTRATION
CALL FLOPRE(X) CALCULATE PREDICTED FLOW
CALL FLOW(X) ESTABLISH INFLUENT FLOW PATTERN
CALL AIRSPL(X) OBTAIN AIR FLOWS TO EACH REACTOR
CALL AERSYS(X) GET POWER CONSUMPTION BY BLOWERS
CALL THKNR(X) GET RETURN SLUDGE CONCENTRATION
TUR = CSET(2) GET TURBIDITY MEASUREMENT

CALL DVERK(N,FCN1,X,Y,XEND,TOL,IND,C,NW,W,IER) CHECK THE FLAGS FROM DVERK
IF(IND.LT.0.OR.IER.GT.0)GOTO 23

DO 66 K=1,44 STORE RESULTS IN COMMON
   66 RES(K)=Y(K)

CALL GETTSS(X) CALCULATE TSS IN EACH REACTOR
CALL WASTE(X) CALL WASTE FLOW CONTROLLER
CALL SHRHHIS(X) UPDATE SHORT HISTORY FILE
CALL INTEGR(X) START OUTPUT LAST DAY

LBEG = LEND - 96 OUTPUT LAST 24 HOURS OF SIMULATION
IF(L.LT.LBEG)GOTO 22
XX = X - LBEG*0.25
CALL OTP(N,XX,Y)

TPWR = HPOWER(1)+HPOWER(2)+HPOWER(3)+HPOWER(4)
SPWR = TPWR/4. + SPWR

CONTINUE

WRITE(5,442)SPWR
442 FORMAT(2X,'TOTAL POWER USED = ','F10.1,' KWHR ')
STOP

ERROR MESSAGE

CONTINUE
WRITE(5,241) IND,IER
241 FORMAT('PROBLEMS WITH DVERK IND=',15,' IER=',15)
SUBROUTINE INI(N,Y)

CALLED FROM : MAIN PROGRAM
OTHER SUBS REQUIRED : NONE

THIS SUBROUTINE SETS UP INITIAL CONDITIONS FOR ALL
THE VARIABLES.

N       V(N),V(N+11),V(N+22),V(N+33)

1       SD  - SOLUBLE SUBSTRATE
2       XP  - STORED PARTICULATE SUBSTRATE
3       XS  - STORED MASS
4       XA  - ACTIVE MASS
5       XI  - INERT MASS
6       XNS - NITROSOMONAS CONCENTRATION
7       XNB - NITROBACTER CONCENTRATION
8       SNH3 - AMMONIA CONCENTRATION
9       SNO2 - NITRITE CONCENTRATION
10      SNO3 - NITRATE CONCENTRATION
11      DO  - DISSOLVED OXYGEN

MODEL ASSUMES FOUR C*-S IN SERIES

COMMON /PATERN/ FOLD,CEIL
COMMON /CONTRL/ DOX,SETP,SCOUR,AIRF,INTEG,DER

DO 5 I = 1,4
L = (I-1)*11
Y(L+1) = 5.0
Y(L+2) = 2.0
Y(L+3) = 10.0
Y(L+4) = 600.0
Y(L+5) = 1800.0
Y(L+6) = 30.0
Y(L+7) = 30.0
Y(L+8) = 10.0
Y(L+9) = 1.0
Y(L+10)= 30.0
Y(L+11)= 3.0

FOLD(1)= 200.0
FOLD(2)= 100.0
FOLD(3)= 150.0
FOLD(4)= 0.0

CEIL(1) = 9.0
CEIL(2) = 8.0
CEIL(3) = 8.0
CEIL(4) = 6.0

C
AIRF(1) = 3800.0
AIRF(2) = 2600.0
AIRF(3) = 2000.0
AIRF(4) = 2000.0

C
RETURN
END
SUBROUTINE FCN1(N,X,Y,YPR)

CALLED FROM : DVERK (IMSL ROUTINE)

SUBROUTINES CALLED : NONE

COMMUNICATION WITH OTHER SUBROUTINES IS ACHIEVED
THROUGH COMMON AREAS IN MEMORY

THIS SUBROUTINE CONTAINS EQUATIONS FOR THE BIOLOGICAL
REACTOR. VARIABLES ARE IN VECTOR Y AND DESCRIBED IN
SUBROUTINEINI.

INPUTS :
N - NUMBER OF EQUATIONS (VARIABLES)
X - INDEPENDENT VARIABLE TIME IN HOURS
Y - VECTOR OF VARIABLES AT TIME X

OUTPUTS :
YPR - DERIVATIVES OF Y
Y - VARIABLES VECTOR

PLEASE SEE DOCUMENTATION FOR IMSL ROUTINE DVERK
FOR MORE DETAILED EXPLANATION

COMMON /CONTRL/ DOX,SETP,SCOUR,AIRF,INTEG,DER
COMMON /FLOWS/ FI1,FI2,FI3,FI4,FWA
COMMON /RETURN/ FR,XR
COMMON /OLDPRO/ OLDPRF
COMMON /OXUPT/ CONS
COMMON /PARAM/ RDT,FSHAT,RSD,RH,KSP,RXA,Y1,KFS,KD,
1 OHNS,KDNS,OHNB,KDNB,KN,Y2,FRAC,TNS,YNB,CNIT,CAM,Y11
COMMON /INFLU/ F0,S0,SNH30
COMMON /VECTR/ RES
COMMON /WASTFL/ TSS,FW
COMMON /VOLFLO/ FOR,FMI,V
COMMON /AIR/ KLA
COMMON /PATTERN/ FOLD,CEIL,MAINFL
COMMON /TURBID/ TUR

OXYGEN LIMITATION ON GROWTH

LN =0
DO 55 N=11,44,11
LN=LN+1
DOX(LN)=Y(N)
RS(LN) = RSD*( Y(N)/(Y(N)+0.4) )
RX(LN) = RXA*(V(N)/(V(N)+0.4))

GET RECYCLE RATE

FR = 0.25 * 450.

SET INFLUENT COMPOSITION

SD0 = FRAC*S0
SP0 = (1.0-FRAC)*S0

F01 = FR + FOR(1)

CR1 = (XR/TSS(4))*(FR/V(1))

FS1 = Y(3)/Y(4)
FS1 = FS1/(1.0+FS1)
F1S = Y(3) / (Y(4) + Y(3) + Y(5))
FP1 = Y(2) / (Y(2) + Y(4))

OXYGEN LIMITATION ON GROWTH
OF NITROSOMONAS AND NITROBACTER

MHNS = OHNS * (Y(11)/ (Y(11)+0.5))
MHNB = OHNB * (Y(11)/ (Y(11)+0.5))

SOLUBLE SUBSTRATE

YPR(1) = (FOR(1)/V(1))*SD0 - (FO1/V(1))*Y(1) -
1 (RDT*(FSHAT-F1S) + RS(1))*Y(4)*Y(1)

PARTICULATE SUBSTRATE

YPR(2) = (FOR(1)/V(1))*SP0 - (FO1/V(1))*Y(2) +
1 CR1*Y(35) - RH*Y(4)*Y(2)

STORED MASS

YPR(3) = CR1*Y(36) - (FO1/V(1))*Y(3) +
1 (RH*Y(2) - (RX(1)/Y1)*F1S )* Y(4) +
1 RDT*(FSHAT-F1S)*Y(4)*Y(1)

ACTIVE MASS

YPR(4) = CR1*Y(37) +
1 (RX(1)*F1S - KD - (FO1/V(1)) )*Y(4)
1 + RS(1)*Y1*Y(4)/Y(1)

INERT MASS

YPR(5) = CR1*Y(38) - (FO1/V(1))*Y(5)
1 + Y2*KD*Y(4)

NITROSOMONAS

YPR(6) = CR1*Y(39) + (MHNS*(Y(8)/(CAM+Y(8))))
1 - KDNS - (FO1/V(1)) )*Y(6)

NITROBACTER

YPR(7) = CR1*Y(40) + (MHNB*(Y(9)/(CNIT+ Y(9))))
1 - KDNB - (FO1/V(1)) )* Y(7)

AMMONIA

YPR(8) = (FOR(1)/V(1))*SNH30 + CR1*Y(41)
1 - (FO1/V(1))*Y(8) - KN*RS(1)*Y(4)*Y(1)
1 - (MHNS*Y(8)/(CAM+Y(8)))*YNS)*Y(6) +
1 \times (1.0\times Y2)\times KN\times KD - KN\times RX(1)\times FS1 \times Y(4)

\text{NITRITE}
\begin{align*}
\text{YPR}(9) &= CR1\times Y(42) - (F01/V(1))\times Y(9) \\
1 + (\text{MHNS}\times Y(8) / (\text{CAM} + Y(8)) \times YNS) \times Y(6) \\
1 - (\text{MHN}\times Y(9) / (\text{CNIT} + Y(9)) \times YNB) \times Y(7)
\end{align*}

\text{NITRATE}
\begin{align*}
\text{YPR}(10) &= CR1\times Y(43) - (F01/V(1))\times Y(10) \\
1 + (\text{MHNB}\times Y(9) / (\text{CNIT} + Y(9)) \times YNB) \times Y(7)
\end{align*}

\text{OXYGEN}
\begin{align*}
\text{CONS}(1) &= RX(1)\times FS1 \times (Y1/1)\times Y(4) + RS(1)\times Y1/1\times Y(4)\times Y(1) \\
\text{CONS}(2) &= (1.0\times Y2)\times KD\times Y(4) \\
\text{CONS}(3) &= 3.4\times (\text{MHNS}\times Y(8) / (\text{CAM} + Y(8)) \times Y(6) \\
\text{CONS}(4) &= 1.1\times (\text{MHN}\times Y(9) / (\text{CNIT} + Y(9))) \times Y(7)
\end{align*}

\text{OUR}(1) = \text{CONS}(1) + \text{CONS}(2) + \text{CONS}(3) + \text{CONS}(4)

\begin{align*}
\text{YPR}(11) &= KLA(1) \times (8.5 - Y(11)) - (F01/V(1))\times Y(11) \\
1 - \text{OUR}(1)
\end{align*}

\text{REACTOR 2}

\begin{align*}
\text{FO2} &= \text{FM1}(2) + \text{FOR}(2) \\
\text{FS2} &= Y(14) / Y(15) \\
\text{FS2} &= \text{FS2} / (1.0 + \text{FS2}) \\
\text{F2S} &= Y(14) / (Y(15) + Y(14) + Y(16)) \\
\text{FP2} &= Y(13) / (Y(13) + Y(15))
\end{align*}

\text{SOLUBLE SUBSTRATE}
\begin{align*}
\text{YPR}(12) &= (\text{FM1}(2)/\text{V}(2)) \times Y(1) \\
1 + (\text{FOR}(2)/\text{V}(2)) \times SDO - (\text{FO2}/\text{V}(2)) \times Y(12) - \\
1 \times (\text{RD} \times (\text{FS} \times \text{F2S}) + \text{RS}(1)) \times Y(15) \times Y(12)
\end{align*}

\text{PARTICULATE SUBSTRATE}
\begin{align*}
\text{YPR}(13) &= (\text{FM1}(2)/\text{V}(2)) \times Y(2) \\
1 + (\text{FOR}(2)/\text{V}(2)) \times SPO - (\text{FO2}/\text{V}(2)) \times Y(13) \\
1 - \text{RH} \times Y(15) \times Y(13)
\end{align*}

\text{STORAGE MASS}
\begin{align*}
\text{YPR}(14) &= (\text{FM1}(2) / \text{V}(2)) \times Y(3) - (\text{FO2} / \text{V}(2)) \times Y(14) \\
1 + (\text{RH} \times Y(13) - (\text{RX}(2)/Y1) \times \text{FS2}) \times Y(15) + \\
1 \times \text{RD} \times (\text{FS} \times \text{F2S}) \times Y(15) \times Y(12)
\end{align*}

\text{ACTIVE MASS}
\begin{align*}
\text{YPR}(15) &= (\text{FM1}(2)/\text{V}(2)) \times Y(4) + \\
1 \times (\text{RX}(2) \times \text{FS2} - \text{KD} - (\text{FO2} / \text{V}(2)) \times Y(15) \\
1 + \text{RS} \times Y1 \times Y(15) \times Y(12)
\end{align*}

\text{INERT MASS}
\begin{align*}
\text{YPR}(16) &= (\text{FM1}(2)/\text{V}(2)) \times Y(5) - (\text{FO2} / \text{V}(2)) \times Y(16) \\
1 + Y2 \times KD \times Y(15)
\end{align*}
NITROSOMONAS

\[
\text{YPR}(17) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(6) + \\
1 \text{ ( MHN*S*(Y(19)/(CAM + Y(19))) - } \\
1 \text{ KDN*S - (FO2/V(2)) } * \text{y}(17)
\]

NITROBACTER

\[
\text{YPR}(18) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(7) + \\
1 \text{ ( MHB*N*(Y(20)/(CNIT*Y(20)))) - } \\
1 \text{ KDN*N - (FO2/V(2)) } * \text{y}(18)
\]

AMMONIA

\[
\text{YPR}(19) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(8) + (\text{FO2/V(2)}) \times \text{SNH30} \\
1 - (\text{FO2/V(2)}) \times \text{y}(19) - \text{KN*RS}(2) \times \text{y1*Y(15)} \times \text{y}(12) \\
1 - (\text{MHN*Y}(19)/\text{(CAM + Y(19)}) \times \text{YN} \times \text{y}(17) + \\
1 - (1.0-Y2) \times \text{KN*KD - KN*RX(2)*FS2} \times \text{y}(15)
\]

NITRITE

\[
\text{YPR}(20) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(9) - (\text{FO2/V(2)}) \times \text{y}(20) \\
1 + (\text{MHN*Y}(19)/(\text{CAM + Y(19)})) \times \text{YN} \times \text{y}(17) \\
1 - (\text{MHN*Y}(20)/(\text{CNIT + Y(20)})) \times \text{YN} \times \text{y}(18)
\]

NITRATE

\[
\text{YPR}(21) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(10) - (\text{FO2/V(2)}) \times \text{y}(21) \\
1 + (\text{MHN*Y}(20)/(\text{CNIT + Y(20)})) \times \text{YN} \times \text{y}(18)
\]

OXYGEN

\[
\text{CONS}(5) = \text{RX}(2) \times \text{FS2} \times (Y11/Y1) \times \text{y}(15) \\
1 + \text{RS}(2) \times Y11 \times Y(15) \times Y(12) \\
\text{CONS}(6) = (1.0-Y2) \times KD \times Y(15) \\
\text{CONS}(7) = 3.4 \times (\text{MHN*Y}(19)/(\text{CAM+Y(19)})) \times \text{y}(17) \\
\text{CONS}(8) = 1.1 \times (\text{MHN*Y}(20)/(\text{CNIT+Y(20)})) \times \text{y}(18)
\]

OUR(2) = CONS(5)+CONS(6)+CONS(7)+CONS(8)

\[
\text{YPR}(22) = (\text{FM1}(2)/\text{V}(2)) \times \text{y}(11) - (\text{FO2/V(2)}) \times \text{y}(22) \\
1 - \text{OUR}(2) + \text{KLA}(2) \times (8.5 - \text{y}(22))
\]

**Reactor 3**

\[
\text{FO3} = \text{FM1}(3) + \text{FO3}(3)
\]

\[
\text{FS3} = \text{Y}(25) / \text{Y}(26) \\
\text{FS3} = \text{FS3} / (1.0 + \text{FS3}) \\
\text{F3S} = \text{Y}(25) / (\text{Y}(26) + \text{Y}(25) + \text{Y}(27)) \\
\text{FP3} = \text{Y}(24) / (\text{Y}(24) + \text{Y}(26))
\]

\[
\text{MHNS} = \text{OHNS} \times (\text{Y}(33)/(\text{Y}(33)+0.5)) \\
\text{MHN} = \text{OHNB} \times (\text{Y}(33)/(\text{Y}(33)+0.5))
\]

**Soluble Substrate**

\[
\text{YPR}(23) = (\text{FM1}(3)/\text{V}(3)) \times \text{y}(12) \\
1 + (\text{FO3/V(3)}) \times \text{SD0} - (\text{FO3/V(3)}) \times \text{y}(23) - \\
1 \times \text{RD*T*(FSHAT-F3S)} + \text{RS}(3) \times \text{y}(26) \times \text{y}(23)
\]

**Particulate Substrate**

\[
\text{YPR}(24) = (\text{FM1}(3)/\text{V}(3)) \times \text{y}(13) \\
1 + (\text{FO3/V(3)}) \times \text{SP0} - (\text{FO3/V(3)}) \times \text{y}(24) \\
1 - \text{RH} \times \text{y}(26) \times \text{y}(24)
\]
STORED MASS
\[ YPR(25) = (FML(3)/V(3))*Y(14) - (FO3/V(3))*Y(25) \]
1 + ( RH*Y(24) - (RX(3)/Y1)*FS3 ) * Y(26) +
1 RDT*(FSHAT-F3S)*Y(26)*Y(23)

ACTIVE MASS
\[ YPR(26) = (FML(3)/V(3))*Y(15) + \]
1 ( RX(3)*FS3 - KD - (FO3/V(3)) ) *Y(26)
1 + RS(3)*Y1*Y(26)*Y(23)

INERT MASS
\[ YPR(27) = (FML(3)/V(3))*Y(16) - (FO3/V(3))*Y(27) \]
1 + Y2*KD*Y(26)

NITROSOMONAS
\[ YPR(28) = (FML(3)/V(3))*Y(17) + \]
1 ( MHNS*Y(30)/(CAM + Y(30))) - KDNS
1 - (FO3/V(3)) * Y(28)

NITROBACTER
\[ YPR(29) = (FML(3)/V(3))*Y(18) + \]
1 ( MHNB*Y(31)/(CNIT + Y(31))) - KDNB
1 - (FO3/V(3)) * Y(29)

AMMONIA
\[ YPR(30) = (FML(3)/V(3))*Y(19) + (FOR(3)/V(3))*SNH30 \]
1 - (FO3/V(3))*Y(30) - KN*RS(3)*Y1*Y(26)*Y(23)
1 - ( MHNS*Y(30) / (CAM + Y(30)) ) *YNS )*Y(28) +
1 ( (1.0-Y2)*KN*KD - KN*RX(3)*FS3 ) * Y(26)

NITRITE
\[ YPR(31) = (FML(3)/V(3))*Y(20) - (FO3/V(3))*Y(31) \]
1 + (MHNS*Y(30) / ( CAM + Y(30) ) )*YNS )*Y(28)
1 - ( MHNB*Y(31) / (CNIT + Y(31) ) )*YNB )* Y(29)

NITRATE
\[ YPR(32) = (FML(3)/V(3))*Y(21) - (FO3/V(3))*Y(32) \]
1 + ( MHNB*Y(31) / (CNIT + Y(31) ) )*YNB )*Y(29)

OXYGEN
\[ CONS(9) = RX(3) * FS3 * (Y11/Y1)*Y(26) \]
1 + RS(3) * Y11 * Y(23) * Y(26)
CONS(10) = (1.0-Y2)*KD*Y(26)
CONS(11) = 3.4*(MHNS*Y(30)/(CAM+Y(30))))*Y(28)
CONS(12) = 1.1*(MHNB*Y(31)/(CNIT+Y(31))))*Y(29)

OURS
\[ \text{OURS}(3) = CONS(9) + CONS(10) + CONS(11) + CONS(12) \]

YPR(33) = CONS(9)*CONS(10)*CONS(11)*CONS(12)
1 - OUR(3) + KLA(3)*(8.5 - Y(33))

REACTOR 4

C C C C C C

FO4 = FML(4) + FOR(4) + FW

FS4 = Y(36) / Y(37)
FS4 = FS4 / (1.0+FS4)
F4S = Y(36) / (Y(37) + Y(36) + Y(38))
FP4 = Y(35) / (Y(35) + Y(37))

C
MHNS = OHNS * (Y(44)/(Y(44)+0.5))
MHNB = OHNB * (Y(44)/(Y(44)+0.5))

C
SOLUBLE SUBSTRATE
YPR(34) = (FML(4)/V(4))*Y(23)
1 + (FOR(4)/V(4))*SD0 - (FO4/V(4))*Y(34)
1 - (RDT*(FSHAT-F4S) + RS(4))*Y(37)*Y(34)

C
PARTICULATE SUBSTRATE
YPR(35) = (FML(4)/V(4))*Y(24)
1 + (FOR(4)/V(4))*SP0 - (FO4/V(4))*Y(35)
1 - RH*Y(37)*Y(35)

C
STORED MASS
YPR(36) = (FML(4)/V(4))*Y(25) - (FO4/V(4))*Y(36)
1 + (RH*Y(35) - (RX(4)/Y1)*FS4) * Y(37)
1 + RDT*(FSHAT-F4S)*Y(37)*Y(34)

C
ACTIVE MASS
YPR(37) = (FML(4)/V(4))*Y(26) +
1 (RX(4)*FS4 - KD - (FO4/V(4))) *Y(37)
1 + RS(4)*Y1*Y(37)*Y(34)

C
INERT MASS
YPR(38) = (FML(4)/V(4))*Y(27) - (FO4/V(4))*Y(38)
1 + Y2*KD*Y(37)

C

C
NITROSOMONAS
YPR(39) = (FML(4)/V(4))*Y(28) +
1 (MHNS*Y(Y(41))/(CAM + Y(41))
1 - KDNS - (FO4/V(4)) ) * Y(39)

C
NITROBACTER
YPR(40) = (FML(4)/V(4))*Y(29) +
1 (MHNB*(Y(42)/(CNIT+Y(42)))
1 - KDNB - (FO4/V(4)) ) * Y(40)

C
AMMONIA
YPR(41) = (FML(4)/V(4))*Y(30) + (FOR(4)/V(4))*SNH30
1 - (FO4/V(4))*Y(41) - KN*RS(4)*Y1*Y(37)*Y(34)
1 - (MHNS*Y(41)/(CAM + Y(41)) )*YNS )*Y(39) +
1 (1.0-Y2)*KN*KD - KN*RX(4)*FS4 ) * Y(37)

C
NITRITE
YPR(42) = (FML(4)/V(4))*Y(31) - (FO4/V(4))*Y(42)
1 + (MHNS*Y(41)/(CAM + Y(41)) )*YNS )*Y(39)
1 - (MHNB*Y(42)/(CNIT + Y(42)) )*YNB )*Y(40)

C
NITRATE
YPR(43) = (FML(4)/V(4))*Y(32) - (FO4/V(4))*Y(43)
1 + (MHNB*Y(42)/(CNIT + Y(42)) )*YNB )*Y(40)

C
OXYGEN
CONS(13) = RX(4)*FS4*(Y11/Y1)*Y(37)
1 + RS(4)*Y11*Y(37)*Y(34)
CONS(14) = (1.0-Y2)*KD*Y(37)
CONS(15) = 3.4*(MHNS*Y(41)/(CAM+Y(41)))*Y(39)
CONS(16) = 1.1*(MHNB*Y(42)/(CNIT+Y(42)))*Y(40)
C

OUR(4) = 3.4*( MHNS* Y(41)/( CAM + Y(41) ))*Y(39)
1 + 1.1*(MHNB*Y(42)/(CNIT*Y(42)))*Y(40) +
1 ((1.0-Y1)/Y1)*RX(4)*FS4 + (1.0-Y2)*KD )*Y(37)

C

YPR(44) = (FML(4)/V(4))*Y(33) - (FO4/V(4))*Y(44) -
1 3.4*( MHNS* Y(41)/( CAM + Y(41) ))*Y(39) -
1 1.1*(MHNB*Y(42)/(CNIT*Y(42)))*Y(40)
1 + KLA(4)*(8.5 - Y(44)) -
1 (((1.0-Y1)/Y1)*RX(4)*FS4 + (1.0-Y2)*KD )*Y(37)

C

DO 11 J=1,4
11 SCOUR(J)=1000.0*OUR(J)/TSS(J)

C

FI1 = FOR(1)
FI2 = FOR(2)
FI3 = FOR(3)
FI4 = FOR(4)
FWA = FW

C

RETURN
END
SUBROUTINE PARSUB (X)

CALLED FROM : MAIN PROGRAM
CALLING OTHER SUBROUTINES : NONE

THIS SUB WILL SUPPLY THE PARAMETER VECTOR TO THE MAIN PROGRAM CONTAINING EQUATIONS FOR ACTIVATED SLUDGE PROCESS

NOTATION AS IN STENSTROM'S AND CLIFFT'S THESIS

REAL MHNS, MHN0, KD, KDNS, KDN0, KN, KFS, KSP
COMMON /PARAM/ RDT, FSHAT, RSD, RH, KSP, RXA, Y1, KFS, KD,
     1 OHNS, KDNS, OHN0, KDN0, KN, Y2, FRAC, YNS, YNB, CNIT, CAM, Y1I

RDT = 0.015
FSHAT = 0.45
RSD = 0.0058
RR = 4.0
RH = 0.001
KSP = 2.0
RXA = 0.45
Y1 = 0.65
KFS = 0.2
KD = 0.025
OHNS = 0.02
KDNS = 0.005
OHN0 = 0.04
KDN0 = 0.005
KN = 0.084
Y2 = 0.25
FRAC = 0.7
SDG = FRAC*S0
S00 = (1.0 - FRAC)*S0
YNS = 1.0/0.02
YNB = 1.0/0.02
CNIT = 5.0
CAM = 5.0
Y1I = 1.0 - Y1

RETURN
END
SUBROUTINE INFL(X)

CALLED FROM : MAIN PROGRAM
CALLING OTHER SUBS : NO

THIS SUBROUTINE ESTABLISHES THE VALUES FOR THE
FLOW RATES AND CONCENTRATIONS IN THE INFLUENT TO
THE PLANT

SIGNAL IS A SINE WAVE WITH A PERIOD OF 24 HOURS

REAL F0, S0, X
COMMON /INFLU/ F0, S0, SNH30

F0 = 450.0 - 250.0 * SIN ( 0.26179939 * X)
S0 = 200.0 - 100.0 * SIN ( 0.26179939 * X)
SNH30 = 40.0 - 20.0* SIN ( 0.26179939 * X)

IF (X.GT.157 .AND. X.LE.162.) F0 = F0 + 256.

RETURN
END
SUBROUTINE WASTE(X)

CALLED FROM : MAIN PROGRAM
OTHER SUBROUTINES CALLED : NONE

THIS SUBROUTINE DETERMINES THE WASTE FLOW RATE.
IT CAN BE REPLACED OR SUPPLEMENTED WITH A WASTE
SLUDGE CONTROLLER LATER

REAL X,FO,FW,RES(44),TSS(4)
COMMON /YVECTR/ RES
COMMON /WASTFL/ TSS,FW

SOLIDS = ( TSS(1) + TSS(2) + TSS(3) + TSS(4) )/ 4.  
SET POINT FOR MLSS (AVERAGE)

SPOINT = 4000.

RLIMIT = SPOINT / 2.0

FW = 10.0 + 0.2 * ( TSS(4) - SPOINT )
IF(FW.LT.0.0)FW=0.0
IF(FW.GT.50)FW =50.

610 FORMAT(F12.5)
611 CONTINUE

RETURN
END
SUBROUTINE FLOW(X)

Called from: MAIN PROGRAM
Calling subroutines: SCRCON

This subroutine determines flows needed for the mass balances around the reactor.
It uses scour controller (SCRCON) to determine influent flow pattern for the next period

Inputs:
X - Independent variable time in hours
NR - Reactor index (1,2,3 or 4)
FIN - Total influent flow rate
FWA - Total waste flow rate
FRET - Total return sludge flow rate

Outputs:
F0N - Influent flow rate to reactor NR
FWN - Waste flow rate from reactor NR
FRN - Return sludge flow rate to reactor NR
PN - Flow rate to next reactor
FM1 - Flow into reactor NR from previous reactor
VN - Volume of the reactor NR

REAL F0(4),FM1(4),V(4),CEIL(4),FOLD(4),SETP(4)
REAL DOX(4),SCOUR(4),AIRF(4),INTEG(16),DER(16)
REAL CSET(10)
INTEGER NR,MAINFL
COMMON /CONTRL/DOX,SETP,SCOUR,AIRF,INTEG,DER
COMMON /INFL/FINF,S0,SNH30
COMMON /VOLFLO/F0,FM1,V
COMMON /PATER/ FOLD,CEIL,MAINFL
COMMON /RETURN/ FR,ER
COMMON /ZPROFL/ CSET

Set up volumes:
V(1) = 400.0
V(2) = 700.0
V(3) = 700.0
V(4) = 700.0

CALL SCRCON CONTROLLER
FIN = FINF
CALL SCRCON(X,FIN,F0)

FM1(1) = 0.0
FM1(2) = F0(1) + FR
FM1(3) = FM1(2) + F0(2)
FM1(4) = FM1(3) + F0(3)
RETURN
END
SUBROUTINE OTP(N,X,Y)

CALLED FROM: MAIN PROGRAM
CALLING SUBROUTINES: NONE

THIS SUBROUTINE WRITES THE RESULTS OF THE SIMULATION INTO DATA FILES THAT HAVE TO BE ALLOCATED BEFORE THE PROGRAM CAN BE EXECUTED. COMMAND FILES ALLSIM.CLIST AND ALLOLD.CLIST CAN BE USED TO CREATE AND ALLOCATE ALL THE NECESSARY FILES

REAL Y(N),X,OUR(4),SCOUR(4),DOX(4),AIRF(4)
REAL INTEG(16),SETP(4)
REAL CONS(16),DER(16),CSET(10),HPower(4)
INTEGER IBLO(4)
COMMON /CONTRL/ DOX,SETP,SCOUR,AIRF,INTEG,DER
COMMON /FLOWS/ F1,F2,F3,F4,FWA
COMMON /OXUPT/ CONS
COMMON /PROFL/ CSET
COMMON /BLOPOW/ TDEMAND,HPower,IBLO

DO 1 I=1,11
     K=20+I
     1 WRITE(K,301)X,Y(I),Y(I+11),Y(I+22),Y(I+33)
 WRITE(32,301)X,OUR
 WRITE(33,301)X,SCOUR
 WRITE(34,301)X,AIRF
 WRITE(35,302)X,F1,F2,F3,F4,FWA
 WRITE(40,303)X,CSET
 WRITE(41,301)X,HPower
 DO 2 L=1,4
 I=35+L
    LL = L-1
 2 WRITE(I,301)X,CONS(LL+1),CONS(LL+5),
     CONS(LL+9),CONS(LL+13)

301 FORMAT(5F10.4)
302 FORMAT(6F10.2)
303 FORMAT(F5.2,10F7.0)
RETURN
END
SUBROUTINE AIRSPL(X)

CALLED FROM : MAIN PROGRAM
CALLING SUBROUTINES : AIRFLO

THIS SUBROUTINE CALCULATES KLA FROM AIR FLOW
KLA IN SECOND BASIN IS CALCULATED

REAL KLA(4),X,AIRF(4),DOX(4),OUR(4),SCOUR(4)
REAL DER(16),HPOWER(4),TDEMND,SETP(4),INTEG(16)
INTEGER IBLO(4)
COMMON /CONTRL/DOX,SCOUR,SETP,AIRF,INTEG,DER
COMMON /AIR/ KLA
COMMON /BLOPOW/ TDEMND,HPOWER,IBLO

CALL AIRFLO(X)

DO 1 I=1,4
   KLA(I) = AIRF(I)*0.002904 - 1.3479
1 IF(KLA(I).LE.0)KLA(I)=0.0

RETURN
END
SUBROUTINE AIRFLO(X)

THIS SUB SUPPLIES CURRENT AIR FLOWS FOR CALCULATION OF KLA

REAL X,AIRF(4),DOX(4),OUR(4),SCOUR(4)
REAL INTEG(16),DER(16),STP(4),IBLO(4)
REAL BIAS(4),PROP(4),PINT(4),HPOWER(4)
COMMON /CONTRL/DOX,STP,SCOUR,AIRF,INTEG,DER
COMMON /BLOPOW/ TDEMND,HPOWER,IBLO

STP(1) = 2.0
STP(2) = 2.0
STP(3) = 2.0
STP(4) = 2.0

CONTROLLER PARAMETERS

PROP(1) = 400.0
PROP(2) = 200.0
PROP(3) = 150.0
PROP(4) = 100.0
PINT(1) = 250.0
PINT(2) = 100.0
PINT(3) = 100.0
PINT(4) = 50.0

DO 11 I=1,4
11 AIRF(I) = AIRF(I) + PROP(I)*(STP(I) - DOX(I))
           + PINT(I)*INTEG(I)

MIXING LIMITS IN ACFM/1000 CU. FT. TANK
BLOIN = 30.

DO 55 J=1,4
C CONVERT TO 1000 CU. FT.
   VOLUM = V(I)*0.0353
   RMIX(J) = BLOIN*VOLUM
C
   DO 22 I=1,4
   IF(AIRF(I).LT.RMIX(I))AIRF(I)=RMIX(I)
   CONTINUE
C
   CHECK TOTAL BLOWER CAPACITY
   TDEMND = AIRF(1) + AIRF(2) + AIRF(3) + AIRF(4)
   IF(TDEMND.LE.18400.)GOTO 31
   DO 33 J=1,4
   AIRF(J) = (AIRF(J)/TDEMND) * 18400.
   TDEMND = AIRF(1) + AIRF(2) + AIRF(3) + AIRF(4)
   CONTINUE
RETURN
END
SUBROUTINE AERSYS(TIM)

C THIS SUBROUTINE CALCULATES THE POWER USED C
C BY BLOWERS OPERATING IN PARALEL. C
C
C CALLED FROM : MAIN C
C
C SUBS USED : SPLITA,POWER C
C
C INPUTS :
C TDEMND - TOTAL AMOUNT OF AIR NEEDED C
C
C OUTPUTS :
C HPOWER(4) - HORSEPOWER USED BY EACH BLOWER C
C IBLO(4) - ON/OFF STATUS OF EACH BLOWER C
C
C REAL TDEMND,BLOWR(4),FLOMAX,SBLOWR(4),HPower(4)
C INTEGER IBLO(4)
C
C COMMON /BLOPOW/ TDEMND,HPOWER,IBLO
C FLOMAX = 4600.
C MAX FLOW FOR A SINGLE BLOWER
C TOTMAX = 4.0 * FLOMAX
C MAX FLOW FOR THE BLOWER HOUSE
C IF(TDEMND.GT.TOTMAX) TDEMND = TOTMAX
C DETERMINE FLOW THROUGH EACH BLOWER
C CALL SPLITA(TIM,FLOMAX,BLOWR)
C GET POWER USED BY EACH POWER
C CALL POWER(BLOWR)

1 CONTINUE

RETURN
END
SUBROUTINE SPLITA(TIM,FLOMAX,BLOWR)
C
C THIS SUBROUTINE CALCULATES THE FLOW THROUGH C
C EACH OF THE FOUR BLOWERS ACCORDING C
C TO THE CONTROL STRATEGY DESCRIBED BY C
C FIGURE IN THE DISSERTATION C
C CALLED FROM : AERSYS C
C CALLING SUBS: NONE C
C
C INPUTS :
C
FLOMAX - MAX FLOW THROUGH THE BLOWER C
TDEMND - TOTAL FLOW NEEDED C

C OUTPUTS :
C
BLOWR(4) - AIR FLOW THROUGH EACH BLOWER C
IBLO(4) - ON/OFF STATUS OF EACH BLOWER C

C
REAL BLOWR(4),TDEMND,FLOMAX,HPOWER(4)
INTEGER IBLO(4)
COMMON /BLOPOW/ TDEMND,HPOWER,IBLO
C
IBLO(2) = 0
IBLO(3) = 0
IBLO(4) = 0
C
SET STARTS AND STOPS
POINT1 = FLOMAX - 0.1*FLOMAX
POINT2 = FLOMAX
POINT3 = 2. * FLOMAX - 0.1*FLOMAX
POINT4 = 2. * FLOMAX
POINT5 = 3. * FLOMAX - 0.1*FLOMAX
POINT6 = 3. * FLOMAX
C
TURN BLOWERS ON
IBLO(1) = 1
IF(TDEMND.GE.POINT2) IBLO(2) = 1
IF(TDEMND.GE.POINT4) IBLO(3) = 1
IF(TDEMND.GE.POINT6) IBLO(4) = 1
C
TURN BLOWERS OFF
IF(TDEMND.LE.POINT1.AND.IBLO(2).EQ.1)IBLO(2)=0
IF(TDEMND.LE.POINT3.AND.IBLO(3).EQ.1)IBLO(3)=0
IF(TDEMND.LE.POINT5.AND.IBLO(4).EQ.1)IBLO(4)=0
C
DETERMINE STATUS
STATUS = IBLO(1) + IBLO(2) + IBLO(3) + IBLO(4)
C
CALCULATE EACH FLOW
BLOWR(1) = IBLO(1) * TDEMND/STATUS
BLOWR(2) = IBLO(2) * TDEMND/STATUS
BLOWR(3) = IBLO(3) * TDEMND/STATUS
BLOWR(4) = IBLO(4) * TDEMND/STATUS
C
RETURN
END
SUBROUTINE POWER(BLOWR)

THIS SUBROUTINE CALCULATES THE AMOUNT OF POWER UTILIZED BY THE BLOWERS. BLOWER CURVE FOR HOFFMAN 75106A CENTRIFUGAL BLOWER IS USED.

CALLED BY: AERSYS

CALLING: NONE

INPUTS:
BLOWR(4)—AIRFLOW IN ACFM THROUGH EACH BLOWER

OUTPUTS:
HPOWER(4) — POWER IN HP FOR EACH BLOWER

REAL BLOWR(4), HPower(4)
INTEGER ISWICH(4)

COMMON /BLOPOW/ TEDMND,HPower,ISWICH

COEFFICIENTS OBTAINED FROM CURVE
A = 2.3572
B = 61.233
C = -6.1435

DO 3 I=1,4
    3 HPower(I) = 0.0

DO 1 I=1,4
    ARFLO = BLOWR(I)/1000.
    HPower(I) = ISWICH(I)*A+B*ARFLO+C*ARFLO*ARFLO

RETURN
END
SUBROUTINE INTEGR(X)

THIS SUBROUTINE OBTAINS ROUGH INTEGRATED VALUES
FOR PERTURBATIONS FROM THE SETPOINT FOR:

DO 1 - 4
SCOUR 5 - 8
NOT ASSIGNED 9 - 16

INTEGRATED VALUES ARE STORED IN COMMON AREA,
IN VECTOR INTEG

REAL Y(44),X,SHORT(52,3),ERR(8,3)
REAL SCOUR(4),DOX(4),AIRF(4),INTEG(16)
REAL DER(16),DOSETP(4)
COMMON /CONTRL/ DOX,DOSETP,SCOUR,AIRF,INTEG,DER
COMMON /SHRTH/ SHORT

SCSETP = 5.0

DO 3 I=1,4
J = I*11

DO 2 L=1,3
ERR(I,L) = DOSETP(I) - SHORT(J,L)
2 ERR(I+4,L) = SCSETP - SHORT(48+I,L)

3 CONTINUE

DO 5 I=1,4
INTEG(I) = (ERR(I,1)+2*ERR(I,2)+ERR(I,3)) /8.0
5 INTEG(I+4) = ( ERR(I+4,1) + 2*ERR(I+4,2)
1 + ERR(I+4,3) ) /8.0

DO 7 I=1,4
7 DER(I) = (ERR(I,2)-ERR(I,1))

RETURN
END
SUBROUTINE GETTSS(I)

C   CALCULATE TSS IN EACH REACTOR
C
REAL TSS(4),RES(44),FW
COMMON /VEXT/ RES
COMMON /WASTFL/ TSS,FW

C
DO 94 K=1,4
   KK = (K-1)*11
   TSS(K) = RES(2+KK) + RES(3+KK) + RES(4+KK)
   1 + RES(5+KK) + RES(6+KK) + RES(7+KK)

C
CNV1 = 1.5
CNV2 = 0.7
CNV3 = 0.66

C
DO 95 K=1,4
   TSS(K) = TSS(K)/(CNV1*CNV2*CNV3)

C
RETURN
END
SUBROUTINE SHRHS(X)

THIS SUB UPDATES THE COMMON AREA SHRTH
VALUES ARE KEPT FOR RES VECTOR, SCOUR AND
TSS ARRAYS.

STRUCTURE OF THE SHORT HISTORY DATA BASE IS:

RES       1-44
SCOUR     45-48
TSS       49-52

REAL DOX(4),AIRF(4),SCOUR(4),INTEG(16)
REAL DER(16),DOSETP(4),CEIL(4)
REAL RES(44),SHORT(52,3),TSS(4),FW,FOLD(4)

COMMON /WASTFL/ TSS,FW
COMMON /YVECTR/ RES
COMMON /SHRTH/ SHORT
COMMON /CONTRL/ DOX,DOSETP,SCOUR,AIRF,INTEG,DER
COMMON /PATTERN/ FOLD,CEIL,MAINFL
COMMON /FLOWS/ F01,F02,F03,F04,FWA

STORING VECTOR RES

DO 1 I=1,44
           SHORT(I,3) = SHORT(I,2)
 1           SHORT(I,2) = SHORT(I,1)
           SHORT(I,1) = RES(I)

SLIDING THE REST OF FILE

DO 2 I=45,52,1
           SHORT(I,3) = SHORT(I,2)
 2           SHORT(I,2) = SHORT(I,1)

UPDATING SCOUR AND TSS

DO 3 I=1,4
           SHORT(44+I,1) = SCOUR(I)
 3           SHORT(48+I,1) = TSS(I)

FOLD(1) = F01
FOLD(2) = F02
FOLD(3) = F03
FOLD(4) = F04

RETURN
END
SUBROUTINE FLOPRE(X)

C OBTAIN FLOW PREDICTION.
C TO MAKE THINGS SIMPLER AND CHEAPER, THE IDEAL FLOW
C PREDICTION (ACTUAL FUTURE VALUE) IS OBTAINED.
C THIS MAY BE SUBSTITUTED WITH TIME SERIES ALGORITHM.

REAL X,XHAT,FHAT,DOX(4),AIRF(4),INTEG(16)
REAL DO(4,3),SETP(4),DER(16),SHORT(52,3)
REAL DOHAT(4)

COMMON /CONTRL/DOX,SETP,SCOUR,AIRF,INTEG,DER
COMMON /SHRT/ SHORT
COMMON /PREDIC/ FHAT,DOHAT

C

R = X + 0.25
FHAT = 450.0 - 250.* SIN(0.26179939*R)
1 + 50.0*SIN(5.*(X+0.5))

RETURN
END
SUBROUTINE SCRCON(X,FIN,F0)

C THIS SUBROUTINE DETERMINES THE CONTACTING PATTERN
C FROM THE SCOUR PROFILE

REAL X,F0(4),SCOUR(4),AIRF(4),INTEG(16)
REAL OFFS(4),CEIL(4),FALL(4),FOLD(4),DOHAT(4)
REAL DER(16),DOX(4),SETP(4),CSET(10)
COMMON /CTRL/DOX,SETP,SCOUR,AIRF,INTEG,DER
COMMON /PREDIC/ FTHAT,DOHAT
COMMON /PATTERN/ FOLD,CEIL,MAINFL
COMMON /XPROFL/ CSET
COMMON /RETURN/FR,XR

C PROVIDE NO CONTROL OPTION
C SET ICONTR TO ENABLE CONTROL

ICONTR = 1
IF(ICONTR.EQ.0)GOTO 506
FR = 0.25 * 450.
C NORMAL RECYCLE FLOW RATE
C ENABLE/DISABLE SPECIAL PROCEDURE
ISPECL = 0
IF(ISPECL.EQ.0)GOTO 508

IF(CSET(2).GE.20.) XC = X + 1.5
C CHECK THE TURBIDIMETER IN LAYER 2
C CHECK DURATION OF CONTROL
IF(XC.LE.X)GOTO 508
C FIXED PATTERN
WRITE(5,666)
666 FORMAT(1X,' SPECIAL PROCEDURE ACTIVE ')
FR = 0.75 * 450.
F0(1) = 0.0
F0(2) = FIN
F0(3) = 0.0
F0(4) = 0.0
GOTO 507

CZY
C 508 CONTINUE

C GET SCOUR PERTURBATIONS
C
DO 1 I=1,4
1 OFFS(I) = SCOUR(I) - CEIL(I)
C FLOW CHANGE EXPECTATION
OFFT=FTHAT - ( FOLD(1)+FOLD(2)+FOLD(3) )

C SET UP FLAGS
MAINFL = 0
DO 2 I=1,4
IF (OFFS(I) .GE. 0) MAINFL = MAINFL + (2**1) / 2

2 CONTINUE

C CHECK THE FLAGS
C IF (MAINFL .EQ. 0) GOTO 70
IF (MAINFL .EQ. 1) GOTO 71
IF (MAINFL .EQ. 2) GOTO 72
IF (MAINFL .EQ. 3) GOTO 73
IF (MAINFL .EQ. 4) GOTO 74
IF (MAINFL .EQ. 5) GOTO 75
IF (MAINFL .EQ. 6) GOTO 76
IF (MAINFL .EQ. 7) GOTO 77
IF (MAINFL .GT. 7) GOTO 78

FLAG = 0
C IF (OFFT .GT. 7.5) GOTO 91
70 DO 10 J = 1, 3
  FALL(J) = FOLD(J)
10 CEIL(J) = CEIL(J) + 0.5*OFFS(I)

91 CONTINUE
GOTO 80

C FLAG = 1
71 FALL(1) = FOLD(1) - 70.
FALL(2) = FOLD(2) + 50.
FALL(3) = FOLD(3) + 20.
GOTO 80

C FLAG = 2
72 FALL(2) = FOLD(2) - 50.
FALL(1) = FOLD(1) + 50.
FALL(3) = FOLD(3)
GOTO 80

C FLAG = 3
73 FALL(1) = FOLD(1) - 50.
FALL(2) = FOLD(2) - 50.
FALL(3) = FOLD(3) + 100.
GOTO 80

C FLAG = 4
74 FALL(3) = FOLD(3) - 50.
FALL(1) = FOLD(1) + 50.
GOTO 80

C FLAG = 5
75 FALL(1) = FOLD(1)
FALL(3) = FOLD(3) - 50.
FALL(2) = FOLD(2) + 50.
GOTO 80

C FLAG = 6
76 FALL(1) = FOLD(1) + 50.
FALL(2) = FOLD(2)
FALL(3) = FOLD(3) - 50.
GOTO 80

C FLAG = 7
77 FALL(1) = FOLD(1) + 50.
FALL(2) = FOLD(2) + 50
FALL(3) = FOLD(3)
C
IF(CEIL(1).EQ.CEIL(2).AND.
   CEIL(2).EQ.CEIL(3))GOTO 98
C
IF(CEIL(3).LT.CEIL(2))CEIL(3) = CEIL(2)
IF(CEIL(2).LT.CEIL(1))CEIL(2) = CEIL(1)
GOTO 99
C
98   DO 17 J = 1, 3
   17  CEIL(J) = CEIL(J) + 1.0
C
99   CONTINUE
   GOTO 80
C
78   FLAG = 8
   FALL(1) = FOLD(1) + 100.
C
80   CONTINUE
C
   DO 5 JJ = 1, 3
      5  IF(FALL(JJ).LT.0)FALL(JJ) = 0.
C
C
   ALLALL = FALL(1) + FALL(2) + FALL(3)
   IF(ALLALL .LE. 0) GOTO 501
   SPLUS = FHTAT - ALLALL
C
C
81   CONTINUE
C
   F0(1) = (FALL(1)/ALLALL) * FIN
   F0(2) = (FALL(2)/ALLALL) * FIN
   F0(3) = (FALL(3)/ALLALL) * FIN
   F0(4) = 0.0
C
   RETURN
C
   NO CONTROL OPTION
C
506  F0(1) = FIN
   F0(2) = 0.0
   F0(3) = 0.0
   F0(4) = 0.0
C
507   CONTINUE
C
   RETURN
C
501  WRITE(5, 425) ALLALL
   425  FORMAT(5X, 'ALL=' , F4.1, ' CHECK AIRFLOW')
C
   RETURN
END
SUBROUTINE THKNR(X)

C THIS SUBROUTINE CALCULATES THE CONCENTRATION OF C
C SOLIDS IN THE UNDERFLOW FROM THE SOLIDS-LIQUID C
C SEPARATOR USING MODIFIED STENSTROM MODEL C
C
C CALLED FROM: MAIN PROGRAM C
C C
C CALLING SUBS : INITLZ, FBLAN, HYD, VELOC C
C C
C INPUTS : PASSED THROUGH INFLU AND WASTFL C
C C
C OUTPUTS : THROUGH COMMON AREA RETURN C
C C
C IMSLS ROUTINE DVERK IS USED C
C C
C REAL CONC(10),VS(10),V0,B,GS(10),GSPUP(10) C
C REAL AREA,VB,VU,FIN,FR,TSS4,TIM,C(24),W(10,9) C
C REAL DCOND(10),TSS(4),CSET(10),GDPDN(10) C
C INTEGER NLYR,N,IBLAN(10)

C COMMON /RTURN/ FRET,XR C
COMMON /INFLU/ F0,S0,SNH30 C
COMMON /WASTFL/ TSS,FW C
COMMON /BRZI/ VS C
COMMON /PARMTR/ AREA,N,DELZ,V0,B C
COMMON /HYDRA/ FIN,FRT,VB,VU C
COMMON /FLBAL/ GS,GSPUP,GSPDN C
COMMON /INPUT/ TSS4 C
COMMON /INITFL/ IFLAG C
COMMON /BLANKT/ IBLAN C
COMMON /XPROFL/ CSET

C

EXTERNAL THICK

C

NLYR = 10
N = 6
TIM = 0.0
TOL = 0.005
IND = 1
NW = 10

C PUT FLOWS IN NEW COMMON AREAS
FIN = F0
FRT = FRET
CZY
IF(X.LT.3.)WRITE(5,333) FRT
333 FORMAT(1X,' IN THKNR FR=',F10.2)
TIM = X
TSS4 = TSS(4)
C INITIALIZE VARIABLES
IF(IFLAG.NE.7)CALL INITLZ(NLYR,CONC)
CALL FBLAN(CONC)

C XEND = TIM + 0.25
C
C CALL HYD(NLYR)
C
C CALL VELOC(NLYR,CONC)
C
C CALL DVERK(NLYR,THICK,TIM,CONC,XEND,
1 TOL,IND,C,NW,W,IER)
C
C XR = CONC(10)
C
C DO 55 I=1,10
55 CSET(I)=CONC(I)
C
RETURN
END
SUBROUTINE INITLZ(NLYR, CONC)

C INITIALIZE VARIABLES FOR THE SLS MODEL
C CALLED FROM : THKNR
C CALLING : NONE
C
REAL CONC(10), VS(10), VO, B, AREA, DELZ
INTEGER NLYR, N
C
COMMON /INITFL/ IFLAG
COMMON /BRZI/ VS
COMMON /PARMTR/ AREA, N, DELZ, VO, B
C
IFLAG = 7
SET Initialization FLAG
C
HEIGHT = 3.66
AREA = 500.
DELZ = HEIGHT/NLYR
S
HEIGHT = 3.66
AREA = 500.
DELZ = HEIGHT/NLYR
C
V0 = 9.19
B = 3.24 E-04
C
DO 1 I = 1, NLYR
1 CONC(I) = I * I * 50.
C
CONC(6) = 5000.
C
RETURN
END
SUBROUTINE HYD(NLAYR)

C HYD CALCULATES BULK FLUID VELOCITIES FOR THE C
C VELOCITY PROFILE DESCRIBED IN DISSERTATION C
C THIS SUB SHOULD BE CHANGED IF DIFFERENT C
C VELOCITY PROFILE IS ASSUMED C
C
C CALLED FROM : THKNR C
C CALLING : NONE C
C
REAL FIN,FRT,VB,VU,AREA
INTEGER NLYR,N

COMMON /HYDRA/ FIN,FRT,VB,VU
COMMON /PARMTR/ AREA,N

VB = (FIN - FRT)/AREA
VU = FRT/AREA

RETURN
END
SUBROUTINE VELOC(NLYR, CONC)

VELOC CALCULATES SETTLING VELOCITIES IN EACH LAYER
V0 AND B ARE DETERMINED FROM BATCH SETTLING TESTS TO FIT THE EXPONENTIAL EXPRESSION: \( VS = V0 \times \exp(-B \times CONC) \)

CALLED FROM : THKNR
CALLING : NONE

REAL CONC(10), VS(10), V0, VB, AREA, DELZ, B
INTEGER NLYR

COMMON /BZII/ VS
COMMON /PARMTR/ AREA, N, DELZ, V0, B

DO 1 I = 1, NLYR
IF(CONC(I).LT.0.0) CONC(I) = 0.0001
IF(CONC(I).GT.35000.) VS(I) = 0.0
IF(VS(I).LT.35000.) VS(I) = V0*EXP(-B*CONC(I))
1 CONTINUE

RETURN
END
SUBROUTINE FLUX(NLYR, CONC)

SOLIDS FLUX IS CALCULATED FOR EACH LAYER
CALLED FROM : THICK
CALLING : NONE

INPUTS THROUGH COMMON BRZI, HYDRA, PARMTR
OUTPUTS THROUGH COMMON AREA FLBAL
NOTATION AS IN DISSERTATION

REAL CONC(10), VS(10), V0, B, GS(10), GSPUP(10)
REAL GSPDN(10), AREA, VB, FIN, FRT, DELZ
INTEGER NLYR, N

COMMON /BRZI/ VS
COMMON /PARMTR/ AREA, N, DELZ, V0, B
COMMON /HYDRA/ FIN, FRT, VB, VU
COMMON /FLBAL/ GS, GSPUP, GSPDN

DO 1 I = N, NLYR
   GS(I) = AREA * VS(I) * CONC(I)
1
DO 2 I = 1, N
   GSPUP(I) = AREA * VB * CONC(I)
2
   GSPDN(I) = AREA * VS(I) * CONC(I)

RETURN
END
SUBROUTINE THICK(NLYR,TIM,CONC,DCOND T)
 C THICK CONTAINS DIFFERENTIAL EQUATIONS FOR
 C FOR THE SLS MODEL OF NLYR NUMBER OF LAYERS
 C
 C CALLED FROM : DVERK
 C CALLING : FLUX
 C
 C INPUTS THROUGH COMMON AREAS BRZI, PARAMTR
 C HYDRA, FLBAL, INPUT
 C PASSED INPUTS : TIM = TIME
 C NLYR = NUMBER OF LAYERS
 C
 C OUTPUTS :
 C IBLAN(10) - FLAGS FOR EXCEEDING 3000 MG/L
 C CONC(10) - SOLIDS PROFILE AT TIME X
 C DCOND T(10) - DERIVATIVE OF PROFILE
 C
 C PLEASE SEE IMSL DOCUMENTATION FOR DVERK
 C
 C REAL CONC(10),DCOND T(10),GS(10),GSUP(10)
 C REAL TSS4,DVOL,DELZ
 C REAL GSPDN(10),VS(10),VU,VB,FIN,FRT,AREA
 C INTEGER NLYR,N,IBLAN(10)

 COMMON /BRZI/ VS
 COMMON /PARAMTR/ AREA,N,DELZ,V0,B
 COMMON /HYDRA/ FIN,FRT,VB,VU
 COMMON /FLBAL/ GS,GSUP,GSUP
 COMMON /INPUT/ TSS4
 COMMON /BLANKT/ IBLAN

 DVOL = AREA * DELZ
 CALL FLUX(NLYR,CONC)

 TOP LAYER
 DCOND T(1) = (GSUP(2) - AMIN1(GSPDN(1),GSPDN(2))
 1 - GSPDN(1))/DVOL
 LAYERS BETWEEN TOP AND FEED
 L = N-1

 DO 1 I=2,L
 CHECK FOR TOP OF BLANKET
 IF(IBLAN(I+1) .EQ. 1 .AND. IBLAN(I) .NE. 1)
 1 DCOND T(I) = ( GSPDN(I-1) -
 1 - AMIN1(GSPDN(I),GSPDN(I+1))
 1 + GSPUP(I+1) - GSPUP(I) )/DVOL
 CHECK FOR BLANKET
 IF(IBLAN(I+1) .EQ. 1 .AND. IBLAN(I) .EQ. 1)
 1 DCOND T(I) = ( AMIN1(GSPDN(I-1),GSPDN(I))
1 - AMIN1(GSPDN(I),GSPDN(I+1))
1 + GSPUP(I+1) - GSPUP(I) )/DVOL
C
  NO LIMIT IF NO BLANKET
C
  IF(IBLAN(I+1).NE.1)DCOND(T) = ( GSPDN(I-1)
  1 - GSPDN(I) + GSPUP(I+1) - GSPUP(I) )/DVOL
C
1 CONTINUE
C
  FEED LAYER
  DCOND(T) = ( (FRT+FIN)*TSS4 +
  1 GSPDN(N-1) - GSPUP(N) -
  1 AREA * VU * CONC(N)
  1 - AMIN1 (GS(N),GS(N+1)) ) /DVOL
C
  K = NLYR-1
J = N+1
C
  DO 2 I=J,K
2   DCOND(T) = ( AMIN1(GS(I-1),GS(I)) -
  1 AMIN1(GS(I),GS(I+1)) - AREA*VU*CONC(I)
  1 + AREA*VU*CONC(I-1) ) /DVOL
C
  BOTTOM LAYER
DCOND(T) = (AMIN1(GS(NLYR-1),GS(NLYR))-
1 AREA*VU*CONC(NLYR)+AREA*VU*CONC(NLYR-1))/DVOL
C
RETURN
END
SUBROUTINE FBLAN(CONC)

SET UP BLANKET FLAGS, LIMIT AT CONCENTRATION OF: 3000 MG/L

CALLED FROM: THKNR CALLING: NONE

REAL CONC(10)
INTEGER IBLAN(10)

COMMON /BLANKT/ IBLAN

DO 1 I=1,10
   IF(CONC(I).GT.3000.)IBLAN(I)=1
   IF(CONC(I).LE.3000.)IBLAN(I)=0
1 CONTINUE

RETURN
END
8.2.2 Pump Station Model

PROGRAM FOR SIMULATING THE PUMP STATION

H - LEVEL IN THE WET WELL
AREA - AREA OF THE WELL (CROSS SECTION)
X1 - ON/OFF SWITCH FOR PUMP 1
X2 - ON/OFF SWITCH FOR PUMP 2
X3 - ON/OFF SWITCH FOR PUMP 3
FLOW1 - PUMP1 FLOW (THOUSANDS OF GPM)
FLOW2 - PUMP2 FLOW (THOUSANDS OF GPM)
FLOW3 - PUMP3 FLOW
QIN - INFLUENT TO THE WET WELL
QOUT - FLOW OUT OF THE WET WELL

ALL THE FLOWS ARE IN CUBIC FEET/HOUR
UNLESS OTHERWISE SPECIFIED

INFLUENT FLOW IS READ FROM UNIT 4
PROGRAM WRITES TO UNIT 1
INPUT FILE : JULY6.DATA
EVERY RECORD OF INPUT FILE CONTAINS
ONE MINUTE DATA IN FOLLOWING FORMAT :
HOUR - I2
MINUTE - I2
FLOW INTO THE WET WELL (QIN) - F10.4
PREDICTED FLOW FOR THE HOUR - F10.4
AVERAGE FLOW FOR THE HOUR - F10.4

FILE NAME : PUMPC.FORT

REAL FLOHAT,HIST(5)
INTEGER TIME(2)
COMMON /LVLS/ H0,H1,H2,H3,H4,H5
COMMON /SNAGA/ POWR1,POWR2,POWR3

HINIT=9.0
FLAG=0.0
FLAG1=0.0
FLAG2=0.0

LEVELS IN WET WELL :
H0 - TURN ALL PUMPS OFF
H1 - TURN ON LEAD PUMP
H2 - TURN ON LAG PUMP
H3 - TURN ON PUMP 3
ALL LEVELS IN FEET

SET UP ON/OFF CONTROL LEVELS
CALL ONOF
TPOW1 = 0.0

CROSS SECTIONAL AREA OF WET WELL
AREA=42.25*11.5*21.
J=0

THIRD PUMP IS ON/OFF, SET FLOW IN 1000G/M
FLOW3 = 7.5
JFLAG=0

DO 100 ITIM=1,1440
IF(ITIM.EQ.1)H=HINIT
READ(4,79,END=44)TIME,QIN,FLOHAT,FLHOUR
FORMAT(2I4,3F10.4)

READING FLOW IN MGD, CONVERT TO CUFT/HOUR
QIN = QIN *1000.0*133.65/24.0

TRY DIFFERENT FLOW RATES
BIGGER = 2.0
QIN = QIN * BIGGER

CALL GETFLO(H,FLAG,FLAG1,FLAG2,
1 FLOW1,FLOW2,SPEED,VALVEP)

TURN BACKUP PUMP ON/OFF
IF(H.GT.13.)X3=1.0
IF(H.LE.4.33)X3=0.
FLOW3 = FLOW3 * X3

CONVERT 1000G/M TO CUFT/HOUR
PUMP1=FLOW1 * 8020.3
PUMP2=FLOW2 * 8020.3
PUMP3=FLOW3 * 8020.3

CALCULATE POWER CONSUMPTION
CALL POWER(H,FLOW1,FLOW2,FLOW3)

CALCULATE TOTAL FLOW OUT
QOUT=PUMP1+ PUMP2+ PUMP3

MASS BALANCE IN WET WELL
DHDT=(QIN-QOUT)/AREA
H = H + DHDT

CONVRT TO 1000 GPM
QIL = QIN / 8020.3
QOT1 = QOUT / 8020.3

C
RTIME = TIME(1)-1. + TIME(2)/60.
IF(RTIME.GT.0.25)
  1 WRITE(1,50)RTIME,H,Q11,FLOW1,FLOW2,POWR1,SPEED
  50 FORMAT(7F10.4)
C
  TPOW1 = TPOW1 + POWR1/60.
C
100 CONTINUE
44 CONTINUE
C
  WRITE(5,111)TPOW1
111 FORMAT(2X,F10.2)
C
  STOP
END
SUBROUTINE GETFLO(LEVEL,FLAG,FLAG1,FLAG2,1,FLOW1,FLOW2,RPM,VALVEP)

C THIS SUBROUTINE TAKES THE LEVEL IN THE
C WET WELL, AND FINDS THE FLOWS THROUGH PUMPS
C USING THE OTHER 5 SUBROUTINES WITH
C CONTROLLERS
C
C INPUT ARGUMENTS:
C  LEVEL - LEVEL IN THE WET WELL (FT)
C  FLAG - RISING OR FALLING LEVEL P #2
C  FLAG1 - RISING OR FALLING LEVEL P #1 (RANGE 1)
C  FLAG2 - RISING OR FALLING LEVEL P #1 (RANGE 2)
C
C OUTPUT ARGUMENTS:
C  FLOW1 - FLOW THROUGH PUMP #1 (1000 GPM)
C  FLOW2 - FLOW THROUGH PUMP #2 (1000 GPM)
C  FLAG - PASSED BACK FOR CONTINUITY
C  FLAG1 - PASSED BACK FOR CONTINUITY
C  FLAG2 - PASSED BACK FOR CONTINUITY
C
C SUBROUTINES:
C  BISECT : FINDS THE POINT WHERE THE PUMP CURVE
C  AND SYSTEM HEAD CURVE MEET BY THE BISECTION
C  METHOD. THIS POINT IS THE PUMP OPERATING
C  POINT, AND GIVES THE FLOW OF THE PUMP
C  GETRPM : FINDS THE SPEED OF PUMP #1 WHICH IS
C  DEPENDENT ON THE LEVEL AND ALSO DEPENDENT
C  ON THE HISTORY OF THE LEVEL
C  GTVALV : FINDS THE VALVE POSITION ON PUMP #2
C  WHICH IS DEPENDENT ON THE LEVEL AND
C  THE HISTORY OF THE LEVEL
C
C VARIABLES:
C  STATH - STATIC HEAD IN FEET
C
C REAL LEVEL
C COMMON /LVLS/ H0,H1,H2,H3,H4,H5
C
C THE STATIC HEAD IS DEPENDENT
C SOLELY ON THE WET WELL LEVEL
C THE BAR SCREEN LEVEL IS 38.5 FEET
C ABOVE THE FLOOR OF THE WET WELL
C
C STATH = 38.5 - LEVEL
C CALL GETRPM(LEVEL,FLAG1,FLAG2,RPM)
C CALL GTVALV( LEVEL, FLAG, VALVEP )
C CALL BISECT(RPM,VALVEP,STATH,1.0,FLOW1)
C CALL BISECT(875.,VALVEP,STATH,2.0,FLOW2)
RETURN
END
SUBROUTINE BISECT(RPM, VALVEP, STAHT, PFLAG, FLOW)

C BISECT TAKES TWO CURVES, THE PUMP CURVE AND THE
C SYSTEM HEAD CURVE, AND FINDS THEIR INTERSECTION
C BY THE METHOD OF BISECTION
C THE INTERSECTION IS THE PUMP OPERATING POINT,
C AND THE ABSCISSA IS THE FLOW THROUGH THE PUMP

C INPUT ARGUMENTS:
C RPM - PUMP IMPELLER SPEED
C VALVEP - THROTTLING VALVE POSITION (% OPEN)
C STAHT - STATIC HEAD IN FEET
C PFLAG - FLAG IDENTIFYING PUMP

C OUTPUT ARGUMENT:
C FLOW - FLOW THROUGH PUMP IN 1000 OF GPM

C SUBROUTINES:
C PCURVE - GIVES PUMPING HEAD FOR GIVEN FLOW
C BY USE OF THE PUMP CURVE
C SCURVE - GIVES SYSTEM HEAD FOR GIVEN FLOW
C BY USE OF THE SYSTEM CURVE

C VARIABLES:
C A - LOWER LIMIT OF ABSCISSA
C B - UPPER LIMIT OF ABSCISSA
C EPS - TOLERANCE FOR FINDING ROOT
C FUNC - FUNCTION FOR WHICH ROOT IS BEING FOUND
C TEMP - TEMPORARY STORAGE OF POINT
C XL - LEFT HAND POINT BEING OBSERVED
C XR - RIGHT HAND POINT BEING OBSERVED

C THE FUNCTION BEING "BISECTED" IS
C THE DIFFERENCE OF THE PUMP
C CURVE AND SYSTEM HEAD CURVE

C IF RPM IS 0 OR VALVE POSITION IS 0,
C THE RESPECTIVE PUMP SHOULD BE OFF

C IF ( PFLAG .EQ. 2.0 ) GOTO 4
IF ( RPM .NE. 0.0 ) GOTO 5
FLOW = 0.0
GOTO 40

4 IF ( VALVEP .NE. 0.0 ) GOTO 5
FLOW = 0.0
GOTO 40

C A = 0 THOUSAND GALLONS PER MINUTE
C B = 8 THOUSAND GALLONS PER MINUTE

5 A = 0.
B = 8.
THE TOLERANCE FOR FINDING THE ROOT IS 0.001
OR 1 GALLON PER MINUTE

EPS = 0.001

XL = A
XR = B

10 CALL PCURVE(XL,RPM,PFLAG,PHEADA)
CALL PCURVE(XR,RPM,PFLAG,PHEADB)
CALL SCURVE(STATH,VALVEP,XL,PFLAG,SYSHA)
CALL SCURVE(STATH,VALVEP,XR,PFLAG,SYSHB)
FUNC = (PHEADA-SYSHA) * (PHEADB-SYSHB)
IF ( FUNC .LT. 0.0 ) GOTO 20
IF ( ( XR - XL ) .LT. ( 2.0 * EPS ) ) GOTO 30
XR = XL
XL = TEMP
XL = ( XL + XR ) / 2.0
GOTO 10

20 IF ( ( XR - XL ) .LT. ( 2.0 * EPS ) ) GOTO 30
TEMP = XL
XL = ( XL + XR ) / 2.0
GOTO 10

30 FLOW = ( XL + XR ) / 2.0
40 RETURN
END
SUBROUTINE PCURVE(FLOW,RPM,PFLAG,PUMPH)

PCURVE CALCULATES A PUMP CURVE
FOR A GIVEN IMPELLER SPEED AND
RETURNS THE PUMPING HEAD FOR A GIVEN FLOW

INPUTS:
FLOW - ABSCISSA CURRENTLY BEING STUDIED (1000 GPM)
RPM - SPEED OF PUMP IMPELLER ( RPM )
PFLAG - DENOTES WHICH PUMP (#1 OR #2)
( PUMP 2 HAS A CONSTANT SPEED )

OUTPUT:
PUMPH - PUMPING HEAD DEVELOPED AT "FLOW"

VARIABLES:
A1 - CONSTANT COEFFICIENT FOR PUMP CURVE FIT
A2 - LINEAR COEFFICIENT FOR PUMP CURVE FIT
A3 - QUADRATIC COEFFICIENT FOR PUMP CURVE FIT
A4 - CUBIC COEFFICIENT FOR PUMP CURVE FIT

IF PUMP # 2 IS BEING OBSERVED, SPEED
IS 875 RPM

IF ( PFLAG .EQ. 2.0 ) RPM = 875.

COEFFICIENTS ARE COMPUTED
WITH A LINEAR RELATIONSHIP TO IMPELLER SPEED

A1 = -24.375 + 0.085 * RPM
A2 = 0.01467 - 0.003946 * RPM
A3 = 0.27333 + 0.0000066442 * RPM
A4 = -0.069399 + 0.00004325 * RPM

THE PUMPING HEAD IS CALCULATED USING
THE COEFFICIENTS IN A CUBIC POLYNOMIAL

SQ = FLOW*FLOW
CUB = FLOW * FLOW * FLOW
PUMPH = A1 + A2*FLOW + A3*SQ + A4*CUB
RETURN
END
SUBROUTINE SCURVE(STATH, VALVEP, FLOW, PFLAG, SYSHED)

SCURVE USES THE SYSTEM HEAD CURVE TO
OBTAIN SYSTEM HEAD FOR A GIVEN FLOW

INPUTS:
STATH - STATIC HEAD IN FEET
VALVEP - THROTTLING VALVE POSITION
FLOW - FLOW FOR WHICH SYSTEM HEAD IS NEEDED

OUTPUT:
SYSHED - SYSTEM HEAD IN FEET

VARIABLES:
COEFF - QUADRATIC COEFFICIENT

IF THE SYSTEM HEAD FOR PUMP 1 IS DESIRED,
THE QUADRATIC COEFFICIENT FOR
THE CURVE FIT IS ALWAYS THE SAME

IF ( PFLAG .EQ. 1.0 ) GOTO 100

FOR PUMP #2, THE SYSTEM HEAD CHANGES
DEPENDING ON THE POSITION
OF THE DISCHARGE VALVE.

THE FOLLOWING SECTION COMPUTES
THE QUADRATIC COEFFICIENT OF THE
CURVE FIT USING INTERPOLATION
BETWEEN KNOWN FRICTION VALUES

IF (VALVEP.LE.100.0.AND.VALVEP.GT.75.0)
  * COEFF=0.1057679+(100.-VALVEP)/25.)*0.02199
IF ( VALVEP .LE. 75.0 .AND. VALVEP .GT. 50.0)
  * COEFF=0.1277553+(75.-VALVEP)/25.)*0.0345782
IF ( VALVEP .LE. 50.0 .AND. VALVEP .GT. 25.0)
  * COEFF=0.1623335+(50.-VALVEP)/25.)*0.1457248
IF ( VALVEP .LE. 25.0 ) COEFF = 0.3080583
GOTO 200

100  COEFF = 0.0832924

200  SYSHED = STATH + COEFF * FLOW * FLOW
RETURN
END
SUBROUTINE GETRPM(LEVEL,FLAG1,FLAG2,RPM)

GETRPM DETERMINES THE RPM FOR THE VARIABLE SPEED PUMP

INPUTS:
LEVEL - WET WELL LEVEL (FEET)
FLAG1 - (CONTROL FUNCTION HAS DOUBLE VALUES IN CERTAIN RANGES.
ONE FUNCTION IS FOR A RISING LEVEL,
THE OTHER IS FOR A FALLING LEVEL)
FLAG1 SHOWS WHETHER LEVEL IS RISING OR FALLING.
FLAG2 - ANOTHER FLAG THAT INDICATES RISING OR FALLING LEVEL IN ANOTHER RANGE

OUTPUT:
RPM - SPEED OF PUMP IMPELLER (RPM)
FLAG1 - FLAG1 IS SENT BACK TO MAIN
FLAG1 = 0. FOR "RISING" LEVEL
FLAG1 = 1. FOR "FALLING" LEVEL
FLAG2 - ALSO SENT BACK TO MAIN
SAME ASSIGNMENTS AS ABOVE

REAL LEVEL
COMMON /LVLs/ H0,H1,H2,H3,H4,H5

SET INITIAL VALUES
EPS = 0.005

LEVEL LESS THAN CUTOFF POINT?
IF ( LEVEL .LE. H1 ) GOTO 900

IF(LEVEL.LE.H2.AND.FLAG2.EQ.0.)RPM=0.
IF(LEVEL.LE.H2.AND.FLAG2.EQ.1.)RPM=700.

IF(LEVEL.GT.H2.AND.LEVEL.LE.H3.AND.FLAG1.EQ.0.)
* RPM=700.+(LEVEL-H2)/(H3-H2)*175.
IF(LEVEL.GT.H2.AND.LEVEL.LE.H3.AND.FLAG1.EQ.1.)
* RPM = 700.

IF(LEVEL.GT.H3.AND.LEVEL.LE.H4.AND.FLAG1.EQ.0.)
* RPM = 875.
IF(LEVEL.GT.H3.AND.LEVEL.LE.H4.AND.FLAG1.EQ.1.)
* GOTO 83
GOTO 91
83 RPM = 700.
WRITE(6,85)LEVEL,FLAG1,RPM
FORMAT(' LEVEL ',F8.4,' ;F1=','F5.2, 'RPM=','F7.2)
* RPM = 700.

91 IF ( LEVEL .GT. H4 .AND. LEVEL .LE. H5 )
* RPM=700.+(LEVEL-H4)/(H5-H4))*175.

IF LEVEL>H5, RUN PUMP AT 100% RATED SPEED
IF ( LEVEL .GT. H5 ) RPM = 875.

SET FLAGS DEPENDING ON THE PRESENT LEVEL
IF ( LEVEL .GT. H4 ) FLAG1 = 1.
IF ( LEVEL .LT. ( H2 + EPS ) ) FLAG1 = 0.
IF ( LEVEL .GT. ( H2 + EPS ) ) FLAG2 = 1.

100 RETURN

IF LEVEL<CUTOFF, SET FLAGS TO ZERO AND SHUT
PUMP OFF

900 FLAG1 = 0.
FLAG2 = 0.
RPM = 0.
GOTO 100
END
SUBROUTINE GTVALV(LEVEL, FLAG, VALVEP)

GTVALV DETERMINES THE VALVE POSITION
FOR THE THROTTLING VALVE ON PUMP #2

INPUTS:
LEVEL - WET WELL LEVEL (FEET)
FLAG - RISING OR FALLING LEVEL

OUTPUT:
VALVEP - POSITION OF THROTTLING VALVE (% OPEN)
FLAG - FLAG IS SENT BACK TO MAIN
FLAG = 0. FOR "RISING" LEVEL
FLAG = 1. FOR "FALLING" LEVEL

H1, H2, H3, H4, H5: "CRITICAL"
LEVELS WHERE STRATEGY CHANGES

REAL LEVEL
COMMON /LVLS/ H0, H1, H2, H3, H4, H5

SET INITIAL VALUES
EPS = 0.04

IF ( LEVEL .LE. H1 ) GOTO 900

IF((LEVEL .LE. H4 .AND. FLAG .EQ. 0.) .AND. VALVEP = 0.
IF((LEVEL .LE. H4 .AND. FLAG .EQ. 1.) .AND. VALVEP = 50.

IF LEVEL IS BETWEEN H4 AND H5,
VALVE "RAMPS" SO THAT IT IS 50 %
OPEN AT H4 AND 100 % OPEN AT H5

IF ( LEVEL .GT. H4 .AND. LEVEL .LE. H5 )
* VALVEP = 50. + ((LEVEL - H4)/(H5 - H4)) * 50.

IF LEVEL IS GREATER THAN H5,
SET VALVE TO 100% OPEN

IF ( LEVEL .GT. H5 ) VALVEP = 100.

SET FLAG AFTER PUMP CUTS ON

IF ( LEVEL .GT. ( H4 + EPS ) ) FLAG = 1.
RETURN

IF LEVEL IS BELOW CUTOFF,
SET FLAG TO ZERO AND SHUT PUMP OFF
900  FLAG = 0.
   VALVEP = 0.
   GOTO 100
   END
SUBROUTINE ONOF

REAL H0, H1, H2, H3, H4, H5
COMMON /LVLS/ H0, H1, H2, H3, H4, H5

ALL PUMPS CUT OFF AT 4.33 FEET
OFFSET = 1.0
H1 = 4.33 + OFFSET
PUMP #1 CUTS ON AT 6.57 FEET
H2 = 6.57 + OFFSET
PUMP #1 AT TOP SPEED AT 7.75 FEET
H3 = 7.75 + OFFSET
PUMP #2 CUTS ON AT 8.16 FEET
(PUMP #1 FALLS TO 700 RPM)
H4 = 8.16 + OFFSET
PUMP #1 AT TOP SPEED AND
PUMP #2 VALVE FULL OPEN AT 11.00 FEET
H5 = 11.00 + OFFSET

RETURN
END
SUBROUTINE POWER(H, FLOW1, FLOW2, FLOW3)

THIS SUB CALCULATES POWER CONSUMED BY EACH PUMP

REAL FLOW1, FLOW2, FLOW3, POWR

COMMON /SNAGA/ POWR1, POWR2, POWR3

MOTOR & DRIVE EFFICIENCY

EFMOTR = 0.9
EFDRIV = 0.95

PUMP EFFICIENCY

STATH = (38.5 - H)*12.
ETAPMP = (26.3316*FLOW1 - 2.0912*FLOW1*FLOW1)/100.

OVERALL EFFICIENCY

ETA = EFMOTR*EFDRIV*ETAPMP

CHANGE TO SI UNITS

G = 1.0
FLO1 = FLOW1*3.785/60.*G
HEAD = STATH * 0.305

IF(ETA.GT.0) POWR1 = (FLO1* HEAD)/ETA
IF(ETA.EQ.0) POWR1 = 0.0

POWR2 = F(FLOW2)

IF(FLOW3.GT.0) POWR3 = FIXED
IF(FLOW3.LE.0) POWR3 = 0.0

POWR = POWR1 + POWR2 + POWR3

RETURN
END