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DYNAMICS AND CONTROL OF SOLIDS-LIQUID SEPARATION IN THE ACTIVATED SLUDGE PROCESS

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DYNAMICS AND CONTROL OF SOLIDS-LIQUID SEPARATION IN THE ACTIVATED SLUDGE PROCESS

by

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February, 1985

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ABSTRACT

The purpose of this investigation was to develop and validate at full-scale dynamic models and control strategies for the activated sludge process capable of predicting both the clarification and thickening functions of the solids-liquid separator. This also included the development of a hydraulic model capable of predicting flow transients through the treatment plant and a solids model to predict MLSS concentrations in the reactors for arbitrary hydraulic forcings. These models were then utilized to derive control strategies to minimize the discharge of effluent suspended solids.

A mixing model was identified using tracer tests. Analysis of the data demonstrates that a tanks-in-series model describes the reactor system better than a dispersion model.

A hydraulic model was developed from mass balances and well known flow equations. The model demonstrates the limited dampening capacity of treatment plants for hydraulic disturbances.

Full scale experiments to identify models and estimate parameters were performed at a 5 MGD wastewater treatment facility in Houston, Texas. A distributed computer monitoring and control system consisting of on-line

instruments, programmable controllers, and a minicomputer were installed at the plant. A table-driven data acquisition and control software package was implemented.

Numerous experiments demonstrated that influent flow rate and pattern were the most important factors affecting the effluent suspended solids concentration at the plant studied. Hydraulic transients had an immediate effect which persisted longer than the actual disturbance. The recycle flow rate had relatively small effects. The sludge blanket level also had little effect until it was very near the water surface. A model was proposed which incorporates these features.

A sludge thickening model developed by Stenstrom was modified to account for the conical bottom of the settler. The model is capable of predicting the return sludge concentration and the accumulation of solids in the settler. Settling parameters were estimated from batch settling tests in a stirred vessel.

These models were utilized to derive an influent pumping strategy to minimize the discharge of effluent suspended solids. The strategy employs flow forecasting and a Simplex optimization routine to utilize the dampening capacity of the wet well in an optimal manner.

A recursive state/parameter estimation technique was adapted for use with the clarification and thickening models. This technique can be used to give better estimates of the model states and update the model parameters from on-line measurements.

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I. INTRODUCTION

The activated sludge process was initially described in a 1914 paper by Arden and Lockett (5), and the first municipal treatment facility in North America was built in 1916 at Houston, Texas. During the 1930's and 1940's, the process became firmly established and was used on a widespread basis. Today, the activated sludge process is the workhorse of modern wastewater treatment. This process is largely responsible for the recent improvements and continual maintenance of our nation's waterways.

Process Description

In the activated sludge process (Figure 1.1), a mixed culture of microorganisms is brought into contact with the wastewater under aerobic conditions. Through various sorption processes, soluble and particulate organic material are removed from the wastewater. The nutrients and organic materials are assimilated by the microorganisms, producing biodegradation products and additional cell mass. After an appropriate contact period, the biomass is separated from the treated liquor by gravity sedimentation. The clear overflow stream is the process effluent which may receive further treatment (such as disinfection) before discharge. The concentrated biomass underflow is returned to the aeration basins to maintain an appropriate microbial concentration. Since the process is a net producer of biomass, some of the microorganisms must be wasted from the system on at least a periodic basis to maintain a "healthy" process with good removal efficiency.

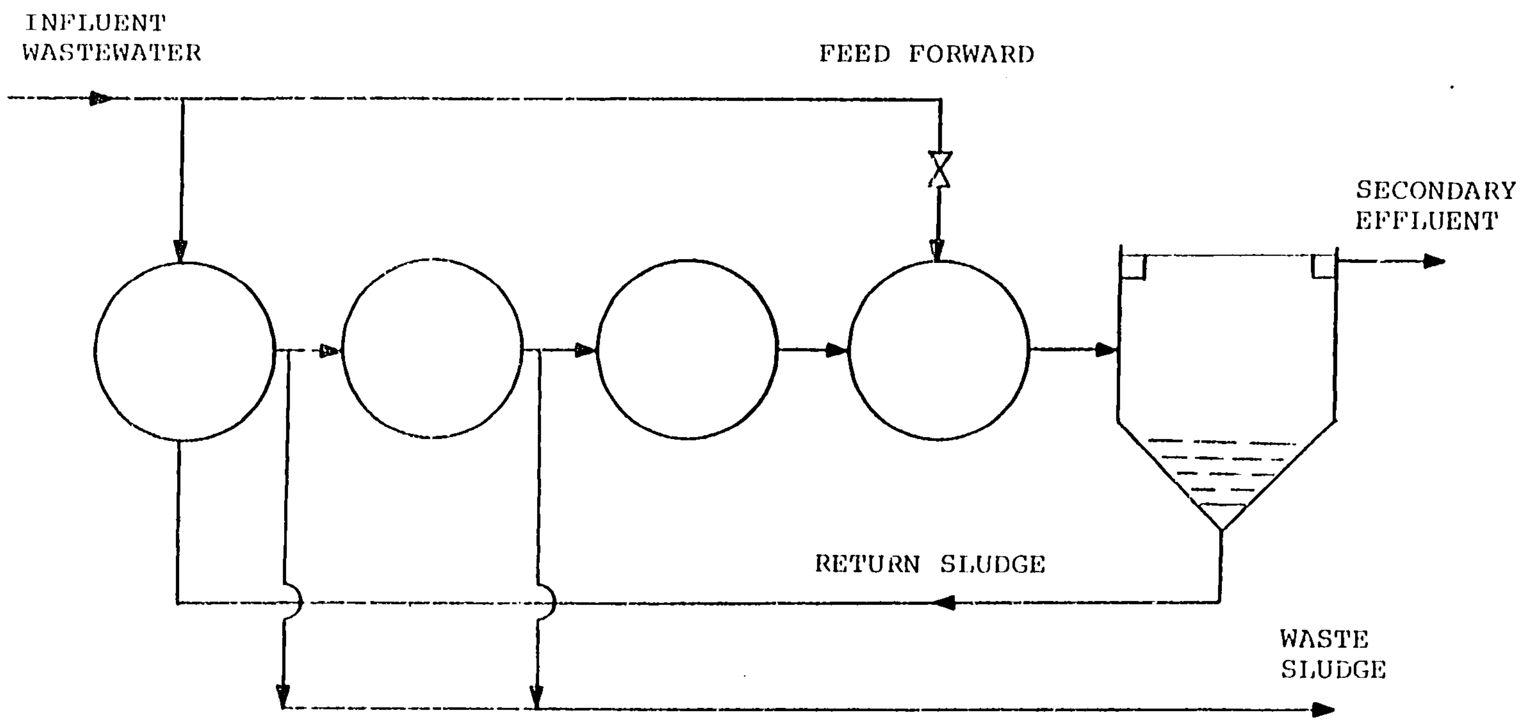

FIGURE 1.1 ACTIVATED SLUDGE PROCESS

The sedimentation basin (or settler) serves several interrelated functions including clarification, thickening, and solids storage. The clarification function is obviously very important since the effluent quality is directly related to the separation of the biological solids from the effluent stream. The thickening capacity of the settler must be sufficient to concentrate the incoming biomass and maintain an appropriate solids concentration in the aeration basins. The volume of the settler must also be sufficient to store solids and prevent them from spilling over the effluent weirs during times of high loading.

Limitations of the Activated Sludge Process

A well designed and operated activated sludge plant is capable of producing a high quality effluent with very low concentrations of dissolved organics. A large portion of the effluent BOD_5 (Biochemical Oxygen Demand) is the result of biological solids not removed during sedimentation rather than unmetabolized influent organics. Every milligram of suspended solids (SS) not removed contributes 0.2-0.7 (43) milligrams of BOD_5, depending on influent composition and operating conditions. A range of 0.5-0.6 mg BOD_5/mg SS (36) is common for domestic sewage and normal operating conditions. This fact suggests that overall improvements in the process will probably take place in the solid-liquid separation phase rather than in improved removal kinetics in the biological reactor. This idea has been noted in a 1979 report of the ASCE Committee on Major Alternatives for Secondary Treatment (27) which states that:

"In properly designed and operated activated sludge systems, influent organics are almost completely degraded with only extremely low residuals appearing in the process effluent. Discharged BOD_5 and SS are virtually completely the result of biological solids in the sedimentation tank effluent. Considering this, attention must be given to operating the oxidation tank and sedimentation tank in ways oriented toward reducing the emission of biological solids in the secondary effluent."

The Activated Sludge Process in Practice

In practice, the performance of the activated sludge process may fall far short of its theoretical capability. In a series of reports to the Congress from 1970 to 1980 (28, 29, 30), the Comptroller General of the United States has documented the poor performance of publicly owned wastewater treatment facilities. The GAO found that at any one time, 50 to 75 percent of the plants were not in compliance with their NPDES (National Pollutant Discharge Elimination System) permits. In a random sampling of 242 plants, the GAO found that 87 percent of the plants had violated their permits for one or more months and that 56 percent had violated their permits for six or more months during the one year study period (1978-79). The GAO report (30) also noted that over twenty-five billion dollars of federal money and several billion dollars of state and local monies had been spent on improvements of wastewater systems through 1980.

Many of the causes of this poor performance record are related to the operation of the plants. In a 1979 survey supported by the U. S. Environmental Protection Agency, Hegg, et al. (63) cited sixty factors limiting treatment plant performance. The highest ranked factors were related to operation and included:

1) a lack of operator application of concepts and testing to process control,
2) a lack of understanding of the wastewater treatment processes,
3) a lack of technical guidance, and
4) a lack of process control testing.

The next six factors were design oriented with the remainder being management policies, financial considerations, equipment malfunctions, and other considerations.

Process Control

Process control is one method of improving the performance of wastewater treatment facilities. Although widely utilized in the petro-chemical industries, process control has not been widely applied to biological treatment processes for a number of reasons. The activated sludge process also presents several difficult control problems including:

1) highly time variant and largely uncontrollable inputs,
2) few manipulatable variables,
3) nonlinear behavior,
4) a wide range of time constants in the interacting systems, and

5) a lack of suitable instruments for measuring variables of interest.

A prerequisite for many forms of advanced process control is an adequate dynamic model of the process. This model should adequately identify the response of the process to arbitrary inputs and predict outputs and internal variables of interest. Dynamic models of biological degradation (16, 23, 42, 110) have recently received a great deal of attention. Dynamic models for the sedimentation process, however, have not received as much attention and have not been as fully developed. Also, there is a deficiency of model validation at full scale.

Research Objectives

It is the purpose of this investigation to develop and identify solids transport models for the activated sludge aeration basins and settler. The models will also be used to develop control strategies to minimize the discharge of effluent suspended solids. Dynamic experiments at a full-scale activated sludge facility will be used for parameter estimation and model discrimination. It is the intent of this investigation to provide models which can be used to provide a better understanding of the interactions of the various wastewater treatment processes and still be simple enough to implement on a small process control computer. The specific objectives of this study are to:

1) develop a reactor mixing model capable of predicting the distribution of solids concentrations in the aeration basins and settler.

2) develop a hydraulic model of the wastewater plant capable of predicting the propagation of hydraulic disturbances through the plant.

3) experimentally verify a clarification model for activated sludge settling and identify the parameters of most importance.

4) experimentally verify the applicability of a sludge thickening model for predicting underflow solids concentrations and the accumulation of solids in the settler.

5) use the models to develop control strategies to minimize the discharge of effluent suspended solids.

6) develop an on-line state/parameter estimation technique to estimate model parameters and states which cannot be easily measured on-line.

II. THEORY AND LITERATURE REVIEW

Process Models

Process models are used widely in all forms of design and automatic process control. Process models may be classified in a great number of ways. One classification of models is dynamic versus steady state. Dynamic models show variations with time while steady state models are time invariant. Since process control deals with time varying functions, dynamic models are used almost exclusively. Steady state models are used primarily in design applications.

Mathematical models may be classified as mechanistic or empirical. Mechanistic models are based upon scientific knowledge about the fundamental biological, chemical, and physical phenomena which govern the process. Models based upon fundamental principles give more insight into process behavior and may be more reliably extrapolated to different operational conditions. Empirical models, on the other hand, rely upon experimental data or mathematical conveniences and cannot always be extrapolated beyond the experimental conditions.

Still another classification of models is as deterministic or stochastic. Deterministic models are those in which the inputs, outputs, and system parameters can be assigned a definite fixed number for a given set of conditions. Stochastic models, however, use statistical techniques to express the model in a mathematical form.

Other common classifications for models include distinctions between linear and nonlinear models, the order of the kinetics, and the number of parameters necessary to describe the model.

The dynamic models to be developed in this research will, whenever possible, be based upon fundamental principles. However, there are a number of areas where theoretical knowledge is lacking, and empirical relationships and expressions have been incorporated. When empirical models are used, the range over which they were derived will be discussed. Therefore, the models developed in this investigation have both mechanistic and empirical components. Investigation of the stochastic components was beyond the scope of this study.

Mixing Models

The mixing characteristics of activated sludge aeration basins vary from close to plug flow to complete mixing. These characteristics are often determined by measuring the output resulting from a pulse input of tracer into the reactor. When plotted as normalized concentration and time, these data are known as C curves. Figure 2.1 shows typical C curves for a plug flow reactor, a completely mixed reactor, and a reactor with arbitrary flow (or mixing characteristics).

In a plug flow reactor, influent streams are instantly mixed together but then move through the reactor with no longitudinal mixing. All fluid elements stay in the reactor for a length of time equal to the theoretical detention time. This type of flow is approximated in tanks with a large length-to-width ratio.

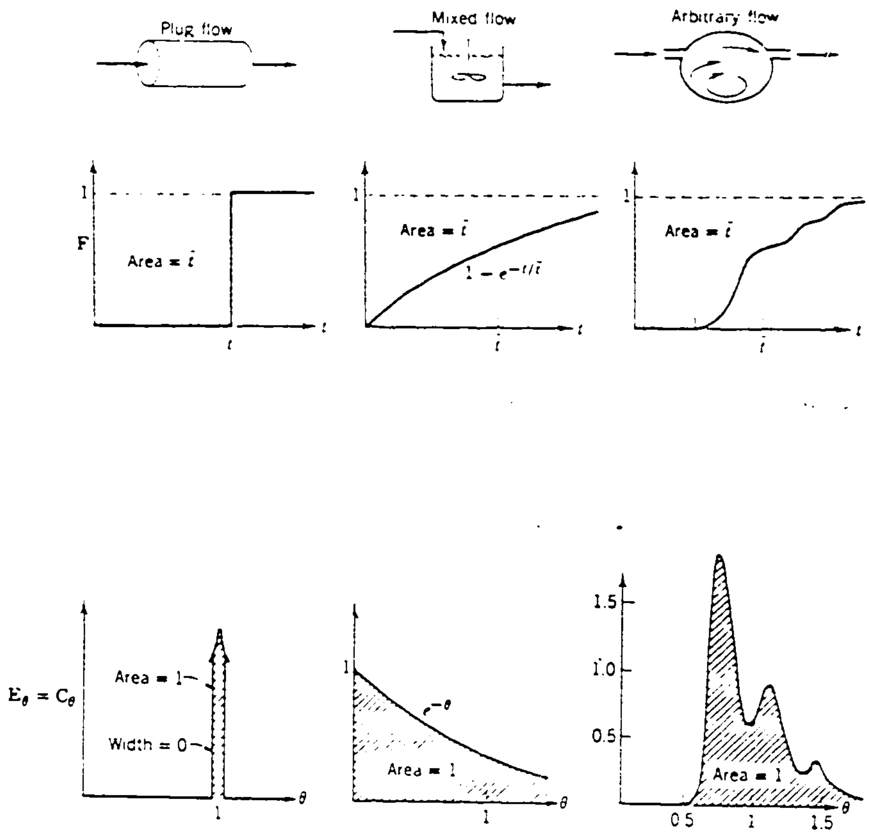

FIGURE 2.1 MIXING MODEL C CURVES

From Levenspiel (74)

In a complete mixing reactor, the influent streams are instantly dispersed throughout the entire reactor volume. Fluid elements leave the reactor in proportion to their statistical population. Complete mixing is approximated in round or square tanks with intense mixing. In practice complete mixing is not difficult to attain and is a satisfactory assumption for many activated sludge plants. Plug flow, however, is more difficult to attain since there is almost always some longitudinal mixing induced by aeration. Many activated sludge plants, therefore, have a mixing condition intermediate between plug flow and complete mixing. Several models are available for this type of arbitrary mixing including the dispersion model, the tanks-in-series model, and various multiparameter models such as the back mix model.

Dispersion Model

The dispersion model is a one parameter model and is based upon a plug flow reactor with longitudinal dispersion. Levenspiel (74) developed the basic differential equation representing this model in dimensionless form as Equation [2.1]. Equation [2.1] must be solved with appropriate boundary conditions for the vessel under study.

$$\frac{\partial C}{\partial \theta} = \left(\frac{D}{u \cdot L}\right) \cdot \frac{\partial^2 C}{\partial Z^2} - \frac{\partial C}{\partial Z} \quad [2.1]$$

where:

C = normalized concentration: Ce/C0,

θ = normalized time: tu/L,

Z = (tu+X)/L,

D = axial dispersion coefficient (L^2/T),

L = length of vessel (L),

u = velocity of flow (L/T),

X = axial distance (L),

t = time (T),

(D/uL) = vessel dispersion number.

The dimensionless group (D/uL) is called the vessel dispersion number and ranges from zero for a plug flow reactor to infinity for a completely mixed reactor. The dispersion model is most appropriate for reactors which vary only a small amount from plug flow (i.e., have small dispersion numbers).

The dispersion number can be determined experimentally from tracer tests by calculating the mean and variance of a pulse input. Relationships between the dispersion number and the mean and variance of the C curve for many boundary conditions are reported by van der Laan (127). Expressions for open and closed vessels are shown as Equations [2.2] and [2.3] respectively.

$$\frac{\sigma^2}{\bar{t}^2} = 2\cdot\left(\frac{D}{u\cdot L}\right) + 8\cdot\left(\frac{D}{u\cdot L}\right)^2 \qquad [2.2]$$

$$\frac{\sigma^2}{\bar{t}^2} = 2\cdot\left(\frac{D}{u\cdot L}\right) - 2\cdot\left(\frac{D}{u\cdot L}\right)^2\cdot(1-\exp(-uL/D)) \qquad [2.3]$$

where:

σ^2 = variance,

$\bar{t}$ = centroid of distribution, mean.

For large extents of dispersion (>0.01) and open vessels, Equation [2.1] can be solved analytically (74) as shown in Equation [2.4]. No analytical expressions are available for other boundary conditions.

$$C(\theta) = \frac{1}{2\cdot\sqrt{\pi\theta(D/uL)}} \exp\left[-\frac{(1-\theta)^2}{4\theta(D/uL)}\right] \qquad [2.4]$$

Tanks-in-Series Model

Another one parameter model widely used in the wastewater field to represent non-ideal flow is the tanks-in-series model. It is particularly appropriate for many activated sludge plants where the aeration basins are physically a series of interconnected basins. In this model, the fluid flows through a series of equally sized ideal stirred tanks. The number of tanks is the one parameter of the model. Again, this parameter may be estimated experimentally from tracer tests. Levenspiel (74) has shown that this parameter is a function of the mean and variance of the C curve as defined in Equation [2.5].

$$1/N = \sigma^2/\bar{t}^2 \qquad [2.5]$$

where:

N = number of tanks in series.

Multiparameter Models

If single parameter models do not adequately account for the mixing patterns observed experimentally, it may be necessary to use a multiparameter model. Many multiparameter models consisting of plug flow, completely mixed, and stagnant regions interconnected in series and/or parallel with various flow patterns have been proposed. However, a model consisting of completely mixed tanks in series with back mixing between adjacent reactors seems particularly applicable to activated sludge facilities. Tuan, et al. (122) have presented response curves for different combinations of parameters to quickly determine the applicability of this model.

The back mix model also has the aesthetically pleasing quality of always resulting in an integer number of reactors.

Hydraulic Models

The calculation of hydraulic profiles through water and wastewater treatment plants is a familiar procedure for many consulting engineers. The concepts for this procedure are taught in undergraduate texts on hydraulics (113) and wastewater treatment (22, 79) as well as several well known reference manuals (35, 57, 134). The purpose of these profiles is to aid in the design of the basins and outflow structures of the facility to maintain an adequate freeboard under all flow conditions. Flow peaking factors for this procedure (6, 55, 60, 142) have been in use at least since 1918.

The same basic ideas used in calculating hydraulic profiles can be extended (with the use of a mass balance) to develop dynamic models capable of predicting hydraulic transients through a treatment plant. Bryant (14) was one of the first to do this. He developed flow equations for a constant volume equalization basin, primary settler, aeration basin, final settler, and chlorination contact tank. Each equation assumed a rectangular weir outlet with free fall conditions as in Equation [2.6].

$$Q = 3.33\ L\ h^{3/2} \tag{2.6}$$

where:

Q = outflow (L^3/T),

L = length of weir (L),

h = depth of liquid over weir (L).

Bryant used a Taylor series expansion to linearize Equation [2.6] and developed a linear differential equation to describe the flow out of each basin. Equation [2.7] is a typical example.

$$\frac{dQ_1}{dt} = \frac{3.33 \cdot L^{2/3} \cdot \hat{Q}^{1/3}}{A} \cdot (Q_0 - Q_1) \tag{2.7}$$

where:

K = unit conversion factor,

$\hat{Q}$ = operating point (L^3/T),

Q_0 = known inflow (L^3/T),

Q_1 = outflow from basin 1 (L^3/T).

Equation [2.7] can be compactly written as:

$$\frac{dQ_1}{dt} = \frac{1}{R \cdot A} \cdot (Q_0 - Q_1) \tag{2.8}$$

where:

R = hydraulic resistance ($L/L^3/T$).

In another report Bryant (15) noted that the linearized equations introduced errors of less than 10 percent for flows that were within 50 percent of the operating point. Although it is not necessary to linearize Equation [2.6] for simulation purposes, Bryant appears to have done this to derive the frequency response of the plant. He concluded that there was no appreciable hydraulic dampening at frequencies of less than 10 cycles per day. Bryant's work is discussed further and compared with the hydraulic model developed in this investigation in a later section.

Measures of Sludge Settling Characteristics

Common measures of activated sludge settling characteristics include the settled volume, the sludge volume index (SVI), the initial settling velocity (ISV), and the effluent suspended solids concentration (Xe).

Settled Volume

The settled volume, expressed in percent, is determined by letting a sample of the activated sludge settle in a one liter graduated cylinder for thirty minutes and reading the interface volume. This same reading is also taken when determining the sludge volume index (SVI). The two tests are related since the settled volume is a measure of the volumetric concentration while the SVI is a specific (per gram) volumetric concentration. Bernard (9), among others, has demonstrated the use of an on-line analyzer for determining the settled volume.

The settled volume is routinely used by many wastewater treatment plant operators to set an appropriate recycle flow rate to obtain a thick return sludge. More recently (1979), Garrett (51) correlated settled volume with initial settling velocity (ISV) or zone settling velocity. He found that the ISV correlated better with settled volume than dry suspended solids concentration. The settling velocity is usually considered a function of dry suspended solids concentration in solids flux theory (39).

Sludge Volume Index

Today the most extensively used SVI is that proposed by Mohlman (81) in 1934. By definition, the SVI is the volume (ml) occupied by one gram of sludge after settling for thirty minutes in a one liter graduated cylinder. More recently, White (137, 138) has proposed a new settling parameter known as the stirred specific volume (SSV) or standard sludge volume index (SSVI). The SSVI is actually the SVI measured in a standard, stirred cylinder. Kalbskopf (68) also reported the German practice of diluting samples such that the settled volume is approximately 200 ml.

The SVI has typically been used as an indicator of thickening characteristics. Slurries with an SVI of less than 100 ml/g tend to settle rapidly and produce thick sludges. Slurries with an SVI greater than 150-200 ml/g tend to settle more slowly and produce light, fluffy sludges.

The SVI has also been related to the clarification characteristics of activated sludge. Summarizing twenty years of operating data, Keefer (69) demonstrated that a higher SVI results in a lower Xe. Figure 2.2 shows some of the results of his study. Fisherstrom, et al. (44) studied twenty Swedish treatment and concluded that the effluent turbidity decreased with higher volumetric sludge volumes. It has been hypothesized that the lower Xe concentrations are a result of improved adsorption of small particulates due to increased surface area of the sludge and increased contact time (slower settling) in the settler.

The SVI has been criticized because it defines only a single point on the settling curve. Hypothetically at least, two sludges with extremely different settling properties could have the same SVI. Dick and Vesilind (38) claim that the SVI does not measure any basic physical property of the sludge. They have criticized the use of the SVI as a research parameter and stated that its only valid application is to monitor changes in sludge characteristics at a single plant. However, there are over fifty years of experience available in interpreting the meaning of the SVI. The test is simple and inexpensive and continues to be routinely used with success as an operational tool in many treatment plants.

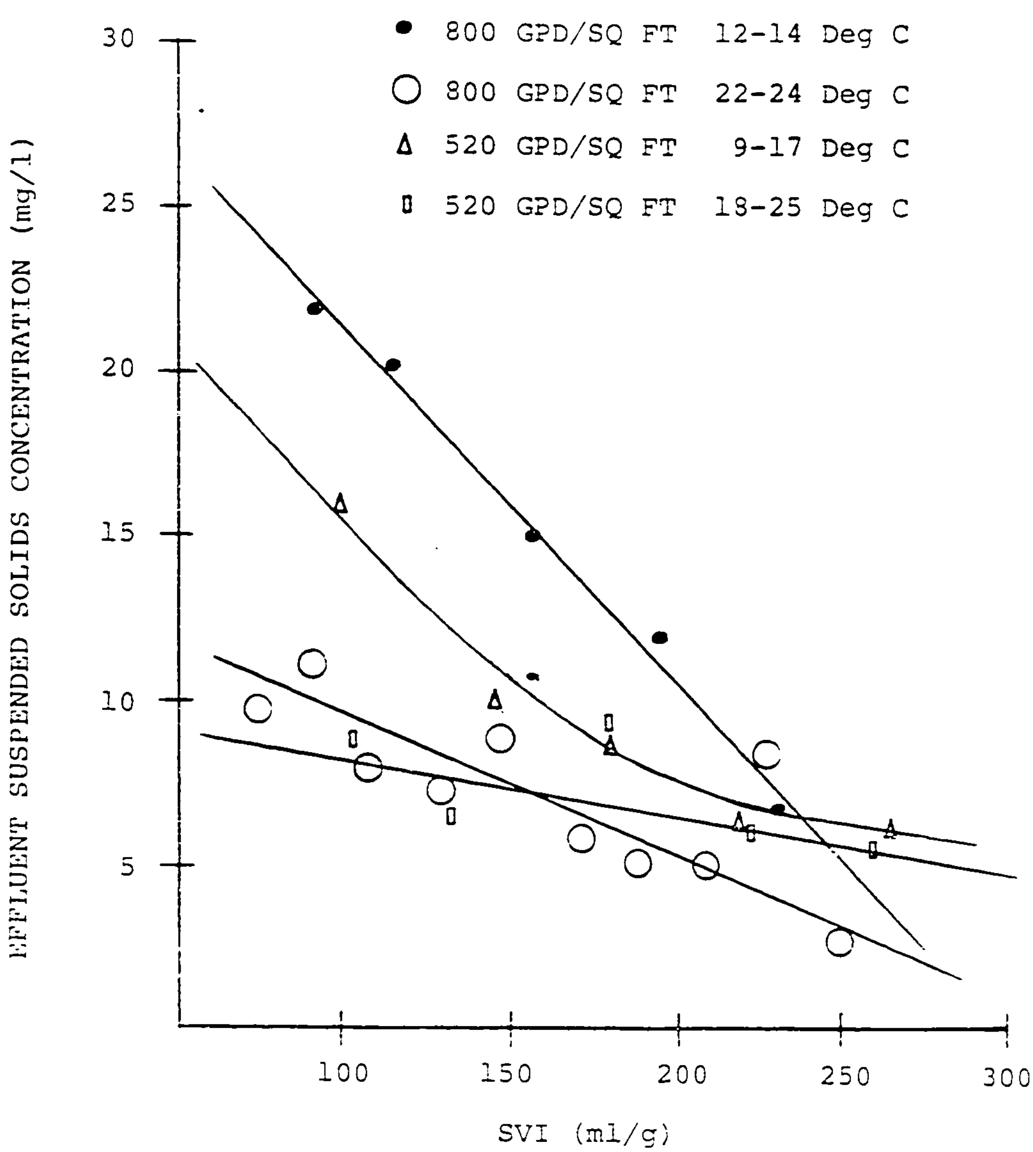

FIGURE 2.2 EFFECTS OF SVI AND TEMPERATURE ON EFFLUENT SUSPENDED SOLIDS

From C. E. Keefer (69)

Initial Settling Velocity

The ISV is a measure of the zone settling velocity of the suspension. It is determined by plotting the solid-liquid interface height versus time for a slurry settling in a column. The ISV is the velocity represented by the first linear portion of this curve. The ISV and the SVI are affected by mixing (37), the column diameter (128) the sludge depth (37) and suspended solids concentration, among other factors. Several instruments (77, 116) for measuring ISV have been available for a number of years.

The ISV is used indirectly as a design parameter in the solids flux methods. The ISV can also be used to directly size the settler. In this technique, the settler is sized such that the overflow rate is always be less than the settling velocity of the influent MLSS. Dick (36) has suggested that this direct use of the ISV is theoretically incorrect and can lead to seriously inaccurate settler design. Wilson and Lee (139), however, have shown that the surface area predicted with the ISV technique is identical to the minimum area predicted by solids flux theory.

Effluent Suspended Solids

A good measure of the settler efficiency and an overall measure of the activated sludge process performance is the effluent suspended solids concentration. Unfortunately, this value is a function both of the settling characteristics of the sludge and the operation of the settler itself. Another indicator of the settling characteristics alone might be the suspended solids concentration in the supernatant after an SVI test. Even this, however, is influenced by the conditions in the settling tube (height, diameter, mixing, etc.) and cannot be considered indicative of the settling characteristics alone (123).

Factors Influencing Settling

A great many factors are known to influence the settling characteristics of activated sludge and the suspended solids concentration (Xe) in the effluent of a secondary settler. These include:

1) physical factors such as reactor and settler dimensions, the inlet well structure, the outlet weir placement, the sludge collection and skimmer mechanisms, mixing intensity, water temperature, and the evolution of gas.
2) hydraulic factors such as the influent and recycle flow rates and hydraulic turbulence caused by flow variations, density currents, thermal gradients, and wind.
3) biological factors such as the influent characteristics including organic content and the absence or presence of nutrients, inhibitors, and toxics.
4) process variables such as the age of the sludge, the MLSS concentration, the dissolved oxygen concentration, the contacting mode, and the sludge blanket height.

The above list is necessarily incomplete and attempts to include only those factors found to be of importance in previous research. An important observation from this previous research is that there is little agreement on how these factors affect the performance of the settler. Conflicting observations have been reported by many researchers.

Reactor Dimensions

The dimensions of the reactors affect the activated sludge settling characteristics by several different mechanisms, usually in conjunction with one or more other variables. These effects are discussed largely under the other variables. It is useful to note, however, that large, well-mixed reactors are less susceptible to shock loads than small ones due to their greater attenuation of concentration disturbances. This property is particularly important when considering shock loads of toxic or inhibitory material.

Settler Dimensions

Camp (17) defined an "ideal settling basin" as a hypothetical settling tank in which settling takes place in exactly the same manner as in a quiescent container of the same depth. Particles are assumed to settle as discrete particles with a constant velocity and are removed from the suspension when they reach the bottom. In this "ideal settling basin," the removal of solids is solely a function of the overflow rate, independent of depth and detention time. Though publicized by Camp, the concept of clarification being a function of overflow rate alone was originated earlier (1904) by Hazen (62). Although the "ideal settling basin" was based upon a number of assumptions which severely limit its direct application, Camp's paper was significant in firmly establishing the overflow rate (hence surface area) as a significant variable in clarification.

Depth and detention time are also significant in clarifier removal efficiency. Fitch (45) conducted tests showing that solids removal is a strong function both of overflow rate and detention time. Chapman (20) demonstrated that side water depth was an important design variable in solids removal. Parker (92) also commented on the importance of deep settlers.

Inlet Well Structure and Baffles

The purpose of the inlet well structure and influent baffles are to reduce the influent flow velocity and to distribute the flow equally across the inlet zone. The extent to which these goals are accomplished is a matter of debate. Fitch (46) designed an inlet well structure which split the influent flow into two oppositely rotating streams. He found that this dissipated the flow energy very effectively and gave good settling. Crosby and Bender (31) experimented with a large number of inlet baffles at full-scale plants and concluded that they had very little effect on performance. They observed that the influent flow settled into a layer that traveled horizontally across the bottom of the settler regardless of the type of baffling.

Sludge Collection and Skimmer Mechanisms

Crosby and Bender (31) observed the effects of the sludge scrapper mechanism. They found that the sludge collectors stirred up a "solids wave" which produced increased effluent solids. At one plant, they found that slowing the sludge scrapper reduced the effluent solids. On the other hand, Chapman (20) in pilot studies found no significant effects of rake speed over a range of 2 to 8 revolutions per hour.

Mixing Intensity

In a study of 24 activated sludge plants in England, Tomlinson and Chambers (119) found that the stirred specific volume (SSV) index could be correlated to the measured dispersion coefficient in the aeration basin. They found that reactors with low dispersion numbers (near plug flow) produced sludges that could be settled more readily (lower SSV). Basins with higher dispersion numbers (near complete mix) produced sludges with poorer settling properties (higher SSV). Heide and Pasveer (64) reported that oxidation ditches operated in the fill-and-draw mode (approximating ideal plug flow), produced sludges with better settling characteristics than those operated continuously.

It is a well established fact that flocculation is a function of mixing intensity (usually expressed as mean velocity gradient). Tuntoolavest, et al. (123) correlated effluent solids with air flow (hence mixing intensity). They found that the effluent solids concentration increased with increasing air flow rates. Parker, et al. (92, 93, 94) have shown that the shear levels in most aeration basins are usually much higher than values found to be optimal for flocculation of activated sludge. They suggested the use of a flocculation chamber between the aeration basins and settler to improve sedimentation. It has been hypothesized that one reason for the improved settling characteristics (lower SVI) of sludges from the UNOX process is the use of tapered mixing. This tapering provides for better flocculation of the sludge before it enters the settler.

Temperature

Temperature affects biological rate constants, oxygen transfer, and the fluid's properties; all of which can affect sludge settling characteristics. As a rule-of-thumb, a 10 degree C rise in temperature will double reaction rates. An increase in temperature will decrease the oxygen saturation concentration and reduce the driving force for oxygenation. Rudolfs and Lacy (101) and Ridenour (100) have shown that settling rates increase with higher temperatures. This is consistent with Stoke's Law (112) of discrete settling in which the settling rate is inversely proportional to the viscosity.

Pflanz (95) presented data from a number of full-scale treatment plants indicating that effluent suspended solids concentration is a strong function of temperature. With similar surface loading rates, effluent solids at 2-3 degrees C was 1.5-2.0 times greater than that at 13-15 degrees C. A plot derived from this data is presented in a following section.

It has been a common observation that differences in temperature between the influent MLSS and the contents of the settler can cause thermal density currents and result in substantial short-circuiting. This short-circuiting can cause a serious deterioration of effluent quality.

In apparent dispute to this common belief, Crosby and Bender (31) measured the inlet and overflow temperatures at eight plants and never found a temperature difference greater than 0.06 degrees C. This does not appear to be sufficient to cause major problems.

Influent and Recycle Flow Rates

General observations indicate that effluent solids increase with the influent flow rate. Virtually all previous research has shown this fact although different researchers have disagreed on the manner (linear or nonlinear). There is still much debate, however, on the influence of recycle flow rate. Pflanz (95) and Sorensen (106) concluded that effluent solids concentration (Xe) was independent of recycle flow rate. Ghobrial (54) and Chapman (20), however, found recycle flow to be as important as influent flow rate. One of the major objectives of this investigation was to clarify these contradictory observations.

Hydraulic Disturbances

Hydraulic disturbances have been shown to have detrimental effects on the removal efficiency of settlers. These disturbances can result from variations of the influent flow, the on/off operation of influent and recycle pumps, and wind.

While working at the Kappala plant in Stockholm, Olsson (87) observed the detrimental effects on a primary settler caused by the on/off operation of influent pumps. In the incident reported, a sudden increase in influent flow rate of approximately 25 percent for 1/2 hour almost doubled the effluent solids. He concluded that there was no effective control action possible to minimize this type of disturbance.

Collins and Crosby (26) proposed the use of a turbulent diffusion model to evaluate the relative amount of mass transport of solids to be expected in the presence of flow variability as compared to that for steady flow. They concluded that flow disturbances have large effects which persist longer than the actual disturbance.

Wind affects the settler by creating an uneven water surface and disrupting the distribution of flow along the effluent weirs. This can result in areas with high local velocities and solids carry-over. Unlevel weir plates can result in a similar situation.

In a theoretical study of population dynamics, Curds (32) studied oscillations of various classes of bacteria and protazoa in the activated sludge process due to diurnal variations in influent flow and concentration. One type of bacteria studied was classified as "dispersed," meaning that they were not removed effectively by the settler. For these bacteria, the hydraulic detention time, rather than solids retention time, determined their relative abundance. This study supports the findings of Cashion, et al. (18) who found that short hydraulic detention times and long sludge ages resulted in the best effluent quality.

Influent Composition

While it is well established that influent composition affects the settling characteristics of activated sludge, quantitative correlations between the two have not been well documented. Qualitatively, however, it has been observed that wastes deficient in nutrients (e.g., nitrogen, phosphorus, iron, etc.) can result in a bulking or poorly settling sludge. Shock loads of toxic materials can give similar results.

Sludge Age

Sludge age (or SRT) has a strong effect on the settling characteristics of activated sludge. Ford and Eckenfelder (49) observed the settling characteristics of activated sludge units treating a domestic and two different industrial wastewaters at various organic loadings over a range of 0.05 to 1.5 lb BOD_5/lb MLSS/day. In each case they observed that the SVI reached a minimum at F/M ratios of 0.15-0.30 ($1/SRT = Y \cdot F/M - Kd$). Higher SVI values were recorded at higher and lower loadings. Bisogni and Lawrence (10) also studied the effects of process loading. They found that the SVI value reached a maximum at sludge ages of 2-3 days. It decreased at both higher and lower values over the range studied (1/2-12 days).

MLSS Concentration

The MLSS concentration entering the settler affects the effluent solids. Different investigators, however, have shown conflicting effects. Pflanz (95), Ghobrial (54), and Chapman (20) concluded that effluent suspended solids concentration (Xe) increased linearly with increasing MLSS. Tuntoolavest, et al. (123) found that Xe was a function of both MLSS and the product of MLSS and air flow rate. Agnew (1) concluded that Xe decreased nonlinearly with increasing MLSS.

Dissolved Oxygen Concentration

The settling characteristics of activated sludge are affected by the dissolved oxygen (DO) concentration in the aeration basins as well as transients in the DO level. Low DO (<0.5 mg/l) is often associated with excess filamentous organisms and bulking sludge. Starkey (108) studied the effects of low DO concentrations and demonstrated that they increased effluent solids. In parallel pilot plants, Wells (135, 136) demonstrated that sludges from the plant with a controlled DO had much better settling characteristics than those from the plant without DO control. It is also a common observation that plants employing anoxic zones (e.g., for denitrification) seldom exhibit bulking.

Sludge Blanket Height

It is suspected that the sludge blanket height can have an effect on the effluent suspended solids concentration (Xe). One hypothesis maintains that operating with a high sludge blanket can decrease Xe by improving the removal of small particulates through adsorption onto other floc particles. If the sludge blanket becomes too high, however, the high velocity zone near the effluent will carry the solids over the weirs.

Little work has been published on the influence of sludge blanket height. Ghobrial (54) found that low blanket heights had little effect while higher blankets increased effluent solids. Crosby and Bender (31), however, noted an opposite effect in which the settlers with high blanket levels produced the clearest effluents.

Models for Clarification

Several of the more important models from the literature are discussed below.

Pflanz's Model

Pflanz (95) observed a number of full-scale treatment plants in Germany to determine what factors affected effluent suspended solids. He proposed that effluent solids concentration was proportional to the product of the overflow rate and the MLSS as in Equation [2.9]. Pflanz called this product the solids surface feed.

$$Xe = K \cdot (F_0/A) \cdot MLSS \quad [2.9]$$

where:

K = model parameter,

F_0 = influent flow (L^3/T),

A = settler surface area (L^2).

A graph derived from data presented in Pflanz's paper is reproduced as Figure 2.3. This figure also shows the pronounced influence of temperature on effluent solids. Pflanz did not attempt to establish a model for temperature effects.

A feature of Pflanz's model is that recycle flow rate does not affect effluent solids since the overflow rate is dependent only on influent flow and settler surface area. Other investigators (15, 54) have not accepted this and have used both the influent and recycle flows in calculating the solids surface feed.

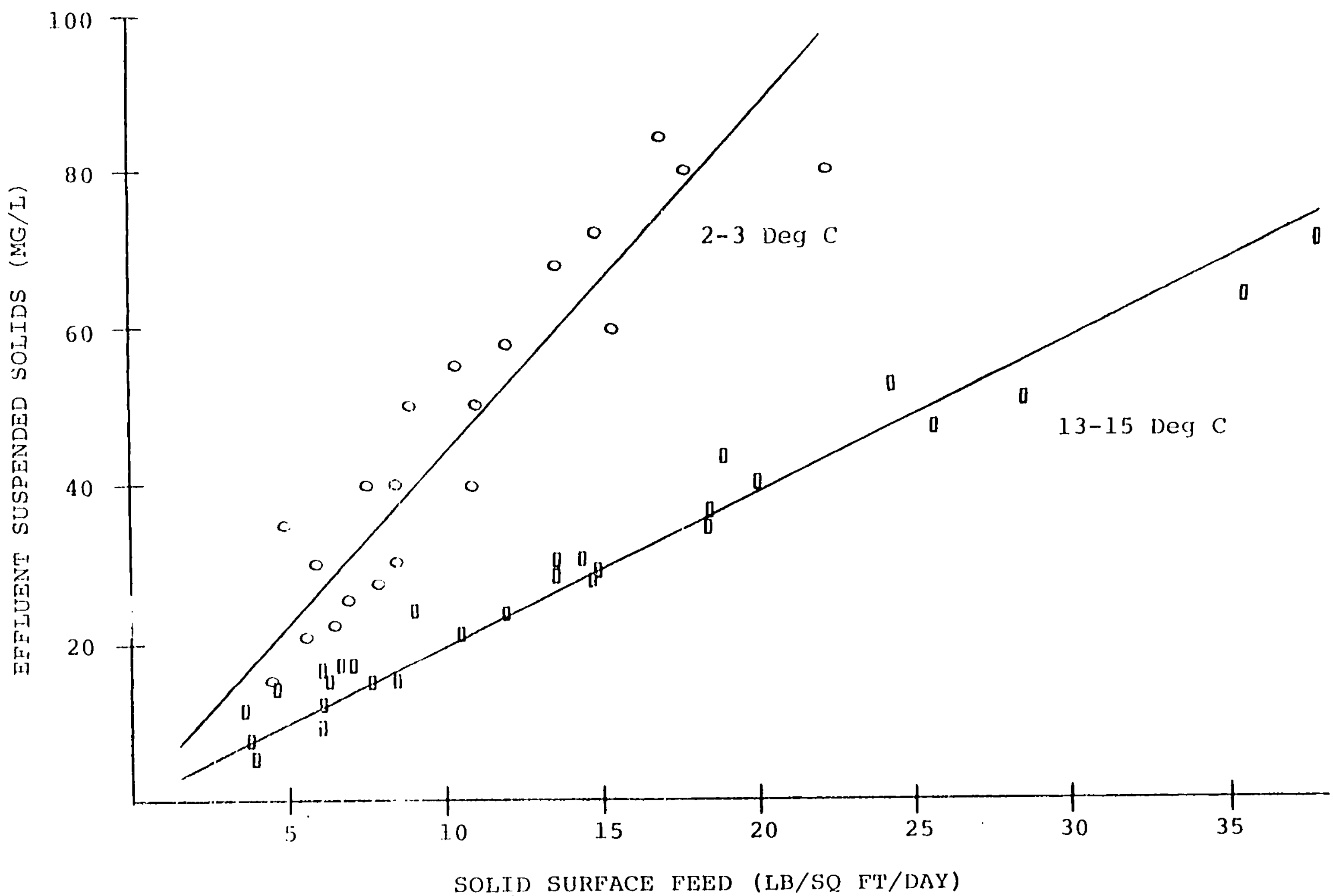

FIGURE 2.3 RELATIONSHIP BETWEEN SUSPENDED SOLIDS AND SOLIDS SURFACE FEED
From Pflanz (95)

Pflanz also observed the effects of SVI variations on Xe. He found that the higher SVI values resulted in higher effluent solids concentrations. This is in conflict with the work of others (31, 69). Parker (92) hypothesized that in the shallow (1.2 to 2.27 m) settling tanks studied by Pflanz, the higher sludge blanket associated with high SVI values interfered with the clarification efficiency.

EPA Model

Working under an Environmental Protection Agency (EPA) grant, Agnew (1) studied three wastewater treatment plants in southern Wisconsin to provide a set of data for formulation of a mathematical model of the final settler. Two models describing effluent solids were proposed. For short periods of time (i.e., 1 month), simple correlations gave very reasonable results. For example, over a 40 day period, Equation [2.10] had a multiple correlation of 0.91.

$$Xe = 18.2 + 0.0136(F_0/A) - 0.0033MLSS \qquad [2.10]$$

Various attempts were also made to incorporate aeration basin parameters such as food to microorganism ratio (F/M) and detention time (DT) but were less successful. Equation [2.11] had a multiple correlation of only 0.63. Predicted and observed effluent solids are shown as Figure 2.4. Note that the original plotting scales have been maintained on this figure.

$$Xe = \frac{382 \cdot (F_0/A)^{0.12} \cdot (F/M)^{0.27}}{(MLSS)^{0.35} \cdot (DT)^{1.03}} \qquad [2.11]$$

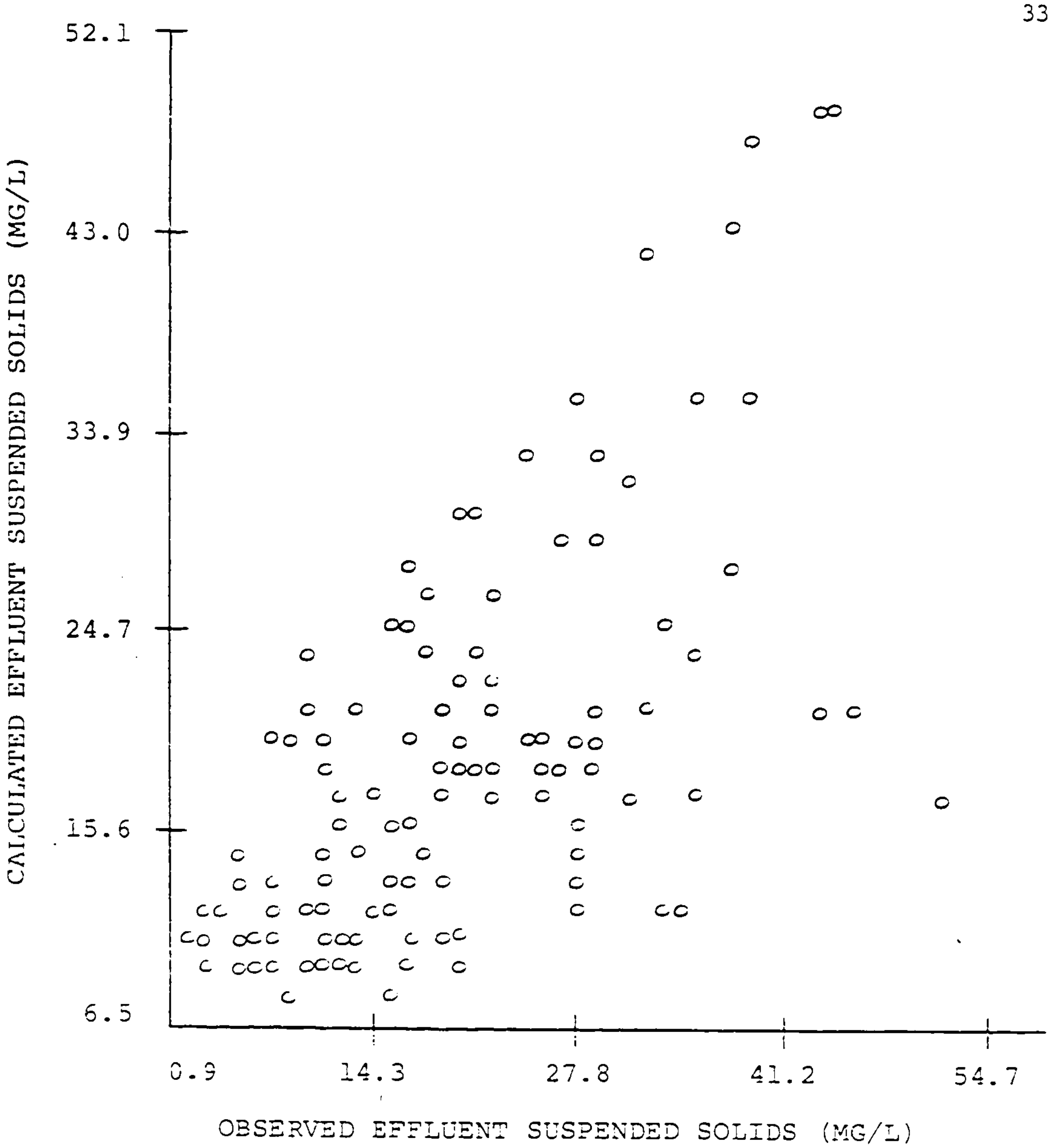

FIGURE 2.4 OBSERVED AND PREDICTED EFFLUENT SUSPENDED SOLIDS CONCENTRATIONS FOR THE EPA MODEL

From Agnew (1)

Equations [2.10] and [2.11] predict a decrease in effluent solids with increasing MLSS. This is exactly the opposite of the behavior predicted by Pflanz's model.

Ghobrial's Model

Ghobrial (54) studied the settling of activated sludge using sludge from a treatment plant in laboratory scale equipment. Ghobrial's experimental design and equipment is discussed further in a later section. He found that the sludge blanket level had a pronounced effect on effluent solids when it exceeded a critical level. In terms of the settled sludge volume and the settler volume, this can be expressed as Equation [2.12].

$$Xe = K_N \cdot \left(\frac{F_0+Fr}{A}\right) \cdot MLSS \cdot \frac{(Vs/V)}{(Vs/V)_{cr}} \qquad [2.12]$$

where:

- K_N = model parameter,
- Fr = recycle flow rate (L^3/T),
- Vs = volume of sludge (L^3),
- V = volume of settler (L^3),
- $(\ )_{cr}$ = critical value of parameter.

Equation [2.12] predicts a deterioration of the settler efficiency as the sludge blanket rises above a critical level. This is in agreement with the observation that solids are scoured from the blanket as it approaches the level of the effluent weirs. However, Ghobrial used a value of 0.12 for (Vs/V)cr which indicates that the lowest effluent solids concentration is obtained by carrying low sludge blanket levels.

When the sludge blanket level was below the critical level, Ghobrial used a Pflanz-like relationship including both influent and recycle flows as in Equation [2.13].

$$Xe = K_N \cdot \left(\frac{F_0 + Fr}{A}\right) \cdot MLSS \qquad [2.13]$$

Tuntoolavest's Model

Tuntoolavest, Miller, and Grady (123) performed a series of steady state tests at pilot scale employing a fractional factorial experimental design. They studied the effects of MLSS concentration, sludge recycle ratio, and air flow rate on effluent solids and sludge thickening characteristics. It should be noted that influent flow rate was maintained constant during each of the tests. Over the range studied, they found these variables had no significant influence on thickening. Their equation for predicting effluent solids, Equation [2.14], had a multiple correlation coefficient of 0.966 for the 14 experiments performed.

$$\begin{aligned} Xe = {} & 24.21 - 0.168(F_{air}) - 1.06(R) + 15.19(MLSS) \\ & - 0.023(F_{air})(MLSS) + 0.005(F_{air})(R) \end{aligned} \qquad [2.14]$$

where:

F_{air} = air flow rate (cu ft/hr),

R = sludge recycle ratio (%),

MLSS = MLSS concentration (g/l).

They concluded that MLSS concentration had the strongest effect, with effluent solids increasing with increasing MLSS. It was also shown that Xe increased with greater air flow rates for recycle ratios of 55 and 80 percent (but not for 30 percent). They concluded that this was evidence of floc breakup.

Chapman's Model

Chapman (20) studied the influence of both process and design variables on effluent solids. He considered seven variables and performed seventeen tests in a factorial design. These experiments were conducted in a modified package plant. He concluded that feed rate (influent and recycle flow), MLSS concentration, and side wall depth all had very significant effects and that the effects of rake speed, feedwell depth, air flow rate, and underflow rate were insignificant. In comparing his work to that of Tuntooiavest, et al. (123), Chapman noted the discrepancy in the effect of air flow rate on effluent solids. He hypothesized that this discrepancy could be attributed to the differences in the size of the test facilities, the aeration equipment, and the influent composition. Equation [2.15] explained 78 percent of the variability of the effluent solids under steady state conditions.

$$Xe = -180.6 + 4.03\,(MLSS) + 133.24\,((F_0+Fr)/A) + SWD\,(90.16 - 62.54\,((F_0+Fr)/A)) \qquad [2.15]$$

where:

SWD = side water depth (L).

It should be noted that Chapman's work supports the notion that both detention time and surface area of the settler are of importance.

Collins and Crosby's Model

Collins and Crosby (26) performed extensive tracer tests on eight full-scale operating clarifiers and developed a clarification model based upon these observations and a number of empirical approximations. They used a turbulent diffusion model to evaluate the relative amount of mass transport of solids to be expected in the presence of flow variability as compared to that of steady flow. This is expressed as Equation [2.16].

$$\frac{M}{\bar{M}} = \frac{1}{P}\int_0^P \left(\frac{V}{\bar{V}}\right)^3 \left(\frac{V_T}{V}\right)^2 dt \qquad [2.16]$$

where:

M = solids transported through settler (M),

$\bar{M}$ = solids transported through settler at a constant velocity V (M),

V = instantaneous velocity (L/T),

$\bar{V}$ = time averaged velocity (L/T),

V_T = velocity characteristic of turbulence level (L/T).

Their model suggests that flow variations act in a nonlinear fashion and result in greater amounts of suspended solids being resuspended and maintained in suspension than would be anticipated for the same steady flow. In general, they concluded that flow disturbances have large effects on effluent solids which tend to persist longer than the flow disturbance. Collins and Crosby, however, give no data to support their model.

Summary of Clarification Models

As discussed in the preceding section, a number of clarification models have been proposed in the literature. Each is based upon a number of assumptions and accounts for various inputs and parameters. Each model has been verified under different conditions. Table 2.1 presents a summary of the models of the preceding sections.

Solids Flux Theory

Due to the economic importance of thickening in many industrial applications, it has been studied for a great number of years starting with the pioneering work of Coe and Clevenger (24) in 1916. Over the years, many refinements to the basic solids flux theory have contributed to its widespread use in design applications. Several of these are mentioned briefly for background material.

Coe and Clevenger realized that settlers could be limited by thickening capacity. They proposed that each concentration in a suspension had a certain capacity (flux) to discharge solids. If a concentration layer in the suspension had a lower solids handling capacity than an overlying layer, it would not be able to discharge solids as fast as they were

TABLE 2.1

SUMMARY OF CLARIFICATION MODELS

Parameter	Model					
	1	2	3	4	5	6
Influent Flow Rate	X	X	X		X	X
Recycle Flow Rate	*		X	X	X	X
MLSS Concentration	X	X	X	X	X	
Activated Sludge Organic Loading		X				
Activated Sludge Detention Time		X				
Sludge Blanket Level			X			
Activated Sludge Air Flow Rate				X	*	
Clarifier Surface Area	X	X	X		X	
Clarifier Side Wall Depth					X	
Flow Rate Variations						X

1 - Pflanz's Model (94), full scale data
2 - EPA Model (1), full scale data
3 - Ghobrial's Model (54), bench scale data
4 - Tuntoolavest's Model (123), bench scale data
5 - Chapman's Model (20), pilot scale data
6 - Collins and Crosby's Model (26), no experimental data

X - Parameter incorporated into model.
* - Found to be insignificant

being received, and the concentration layer would increase in thickness. Similarly, if a layer was able to transmit solids at a faster rate than they were received from the overlying layer, its thickness would remain infinitesimal. For sludges from the activated sludge process, the concentration with the lowest solids capacity (limiting concentration) is usually between the MLSS concentration and the underflow concentration.

Several similar design procedures (37, 61, 129, 141) have evolved from the original flux theory of Coe and Clevenger. These procedures are based upon providing surface area in the settler such that the applied flux is always less than the limiting flux. The limiting flux is usually determined from a series of batch settling tests at various initial concentrations as discussed in a later section. Detailed design procedures are given in several sources (129, 130). The graphical solution proposed by Yoshioka, et al. (141) is particularily useful since different design conditions can be explored merely by striking lines tangent to the settling flux curve.

Kynch (72) proposed that the settling velocity of particles at any point in a suspension is determined only by the local particle concentration. With this assumption, Kynch devised a theory of the propagation of concentration discontinuities by which the entire settling flux curve could be derived from a single batch settling test. Shortly thereafter, Talmage and Fitch (115) used Kynch's theory to devise a simple design procedure which is still presented in many texts (22, 79). However, after many years of use, this procedure has been found to be inapplicable to flocculant materials such as activated sludge (39, 47) since Kynch's main assumption is not strictly true. Recently, Owens (91) has modified the Kynch equation to account for the rising sediment at the bottom of the batch settling column. Unfortunately, these equations are not easily simplified for design purposes.

There is still considerable controversy in determining the "correct" relationship between settling velocity and solids concentration. Vesilind (130) has presented a table (reproduced as Table 2.2) with a partial listing of the various proposed relationships. All are in part empirical.

The equations most commonly used in wastewater applications are the power function curve (25) and the exponential function (128). The exponential relationship was chosen for use in this investigation for two reasons; 1) the equation fit the experimental data well, and 2) at low solids concentrations, the settling velocity approaches a finite constant rather than infinity. This second property is especially important for numerical stability during computer simulations.

Models for Sludge Thickening

Several of the more important sludge thickening models from the literature are discussed below.

Bryant's Model

Bryant (15) developed the first dynamic model for continuous thickening. His model was part of a comprehensive model developed to simulate the

TABLE 2.2

RELATIONSHIPS BETWEEN SOLIDS CONCENTRATION AND SETTLING VELOCITY

Equation	Source
$v = v_0\ (1-KC)^{4.65}$	(99)
$v = v_0\ (1-KC)^2 \cdot 10^{-1.82KC}$	(109)
$v = v_0\ 10^{-aKC}$	(118)
$v = aC^{-b}$	(25)
$v = v_0\ [1 - 2.78\ (KC)^{2/3}]$	(11)
$v = v_0 \exp(-aC)$	(128)
$v = v_0\ [1 + \frac{3}{4} KC\ (1 - \frac{8}{KC-3})]$	(13)
$v = v_0\ (1-KC)^a$	(78)
$v = v_0\ (1-aKC)\ [1 - b(KC)^{1/3}]$	(85)
$v = \dfrac{v_0\ [3 - \frac{9}{2}\ (KC)^{1/3} + \frac{9}{2}\ (KC)^{5/3} - 3\ (KC)^2]}{3 + 2\ (KC)^{5/3}}$	(59)

v = interface velocity of solids concentration, C

v_0 = Stokes settling velocity for a single, discrete particle

K = conversion variable so that KC = volume fraction of solids

a,b = constants (unique to each equation)

After Vesilind (130)

entire activated sludge treatment plant and can be derived by writing a mass balance equation around a differential element in the settler. This mass balance takes the form of a partial differential equation as shown in Equation [2.17].

$$\frac{\partial X}{\partial t} = -[U + \frac{\partial G_s}{\partial X}] \cdot \frac{\partial X}{\partial Z} \qquad [2.17]$$

where:

X = suspended solids concentration (M/L^3),

U = underflow velocity (L/T),

G_s = settling flux ($M/L^2/T$),

Z = vertical distance (L).

Stenstrom (110) noted that a problem with Bryant's formulation was that there was only one equation and two unknowns (Vs and X). Bryant assumed settling velocity to be an algebraic function of solids concentration. An empirical mechanism was employed in the numerical solution technique such that the limiting flux predicted by flux theory was never exceeded.

Stenstrom's Model

Stenstrom (110) further developed Bryant's work and presented a model based upon a one-dimensional continuity equation and solid flux techniques. His model consists of one partial differential equation with boundary conditions imposed at the top and bottom of the settler. This equation is:

$$\frac{\partial X}{\partial t} = -U\cdot\frac{\partial X}{\partial Z} - \frac{\partial(V_s X)}{\partial Z} \qquad [2.18]$$

A number of simplifying assumptions were made in the development of Equation [2.18]. These include:

1) The solids concentration in any horizontal plane is uniform.
2) Mass is conserved over the settler.
3) The settler may be modeled as a plug flow reactor (dispersion is zero).
4) The bottom of the settler represents a physical boundary to sedimentation. Therefore, the settling flux at the bottom of the settler is zero.
5) The settling velocity is a function only of solids concentration except when assumption number six is violated.
6) The settling flux into a differential volume can never 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.

A justification for assumption six may be obtained from an examination of the settling phenomenon between elements. The solids associated with the flux into the ith element must pass through every concentration between X_{i-1} and X_i. Any settling flux over this concentration range could limit the settling flux between these two elements. However, an examination of the settling flux curve shows that there are no local minima over the range of interest. Thus, only the settling fluxes associated with the two concentrations need be checked. A similar argument may be used for the settling flux out of the ith element.

While never experimentally verified, Stenstrom's model gives results which qualitatively agree with observed settler behavior. An important property of the model (induced by assumption six) is the ability to predict "non-ideal" concentration profiles in the settler. Few other models have this ability.

Tracy's Model

Tracy (121) formulated another thickening model based upon Bryant's work and Equation [2.17]. Unlike Bryant, however, he included an upper dilute concentration zone and a completely stirred compression zone. Provisions were made in the computer solution to limit the maximum flux transmitted throughout the thickener to the limiting flux predicted by flux theory.

Tracy's work also represents one of the first attempts to experimentally verify a dynamic model of continuous thickening. Tracy used a laboratory scale thickener and a ferric hydroxide precipitate for floc. His model adequately predicted underflow solids concentration and sludge blanket heights for influent forcings but was less descriptive for underflow forcings. This discrepency was thought to be more of an experimental problem due to the small thickener than an inaccuracy of the model.

Kos' Model

A different approach to modeling the continuous thickening process was developed by Kos and Adrian (71). They proposed that consolidation of solids in a thickener is analogous to transport of mass and momentum in a non-rigid saturated porous media. The general differential equations describing both mass and momentum transport in a porous medium were developed by Raats (98). Kos and Adrian assumed inertia forces were negligible and derived the following differential equation:

$$\frac{d_1}{\rho_s}\frac{\partial \rho_s}{\partial t} = -\frac{\partial}{\partial X}\left[\frac{\rho_1^2}{F}\left(\frac{1}{d_1}\frac{\partial p}{\partial X} - g\right)\right] \qquad [2.19]$$

where:

d_1 = density of liquid (M/L^3),

ρ_s = bulk density of solids (M/L^3),

ρ_1 = bulk density of liquid (M/L^3),

g = gravitational constant (L/T^2),

X = spatial distance (L),

F = resistivity $(M/L^3/T)$,

p = liquid phase pressure $(M/L/T^2)$,

The liquid and solid momentum balances were summed to derive an expression which defines the liquid phase pressure. They discovered that interparticle pressure was a function of the local solids concentration.

$$\partial p/\partial X + \partial \sigma/\partial X = g\,(\rho_1 + \rho_s) \qquad [2.20]$$

where:

σ = effective (interparticle) pressure ($M/L/T^2$).

Kos and Adrian compared computer simulations of the model with laboratory steady state continuous thickening experiments. The model adequately predicted the data within the zone settling portion of the thickener. They presented no simulations or experimental data, however, for dynamic forcings.

George's Model

George (53) developed a thickening model based upon a mass balance on solids, a mass balance on the liquid, and a force balance to accommodate effects of compressive stress. He used both batch and continuous settling tests with a calcium carbonate suspension. The continuity equation for the solids is presented in Equation [2.21]. Both settling velocity (Vs) and depth (Z) are considered positive in the downward direction.

$$\frac{\partial X}{\partial t} = -(V_s + V_w + U)\frac{\partial X}{\partial Z} - X\frac{\partial V_s}{\partial Z} - X\frac{\partial V_w}{\partial Z} \qquad [2.21]$$

where:

V_w = velocity of displaced fluid (L/T).

A mass balance on the liquid over a incremental volume results in Equation [2.22] which defines the change in the upward velocity of displaced liquid with respect to depth.

$$\frac{\partial V_w}{\partial t} = -\frac{U}{X}\frac{\partial X}{\partial Z} - \frac{1}{X}\frac{\partial X}{\partial Z} \qquad [2.22]$$

Using equations developed by Kos and Adrian (71), George simplified the force balance to account for reductions in settling velocity due to compressive forces and mechanical support. His final equation is:

$$V_s = V_0 - f(X) \cdot (\partial X/\partial Z) \qquad [2.23]$$

The function f(X) was empirically approximated with Equation [2.24]. The coefficients of this equation were estimated by a least squares regression of data acquired from batch settling tests.

$$f(X) = K_1 \cdot X^2 + K_2 \cdot e^X + K_3 \qquad [2.24]$$

where:

K_1, K_2, K_3 = experimental coefficients.

George's model presents a fundamental approach to describing the movement of solids within the settler and the influence of displaced fluid and compressive forces on solids transport. However, the basic model still failed to predict a rising sludge blanket during times of overloading. George remarked that this inadequacy was induced by the force balance assumed in the model. He compensated for this by specifying that a thick blanket be formed whenever the limiting flux was reached.

Process Control

Many of the elements of a typical process are shown in Figure 2.5. The process may have many inputs which can be classified either as disturbances or manipulated variables. Manipulated variables are inputs which can be changed by the plant operator or an automatic controller. Disturbances are inputs which cannot be controlled. The system state consists of all variables (e.g., flows and concentrations) of the process. It is usually the intent to control one or more of the process outputs which are therefore called controlled variables. The observed state of the process is usually different than the real system state due to measurement noise.

Process control can be defined as the manipulation of variables to change the state of a process to accomplish some desirable objective. Typical objectives include maintaining a constant output (regulator problem), having an output follow a predetermined or changing path (servo problem), and minimizing or maximizing some quantity (optimization).

Process control systems can be divided into a number of categories with the two most common being feedback and feedforward. Figure 2.6 shows the characteristics of a typical feedback control system. Controlled outputs are sensed, and the measured output is compared to the set point. Any deviation between the set point and the measured quantity is used to manipulate the inputs to decrease the error. Since it is necessary to disturb the controlled variable before any control action is initiated, feedback controllers can never achieve perfect control.

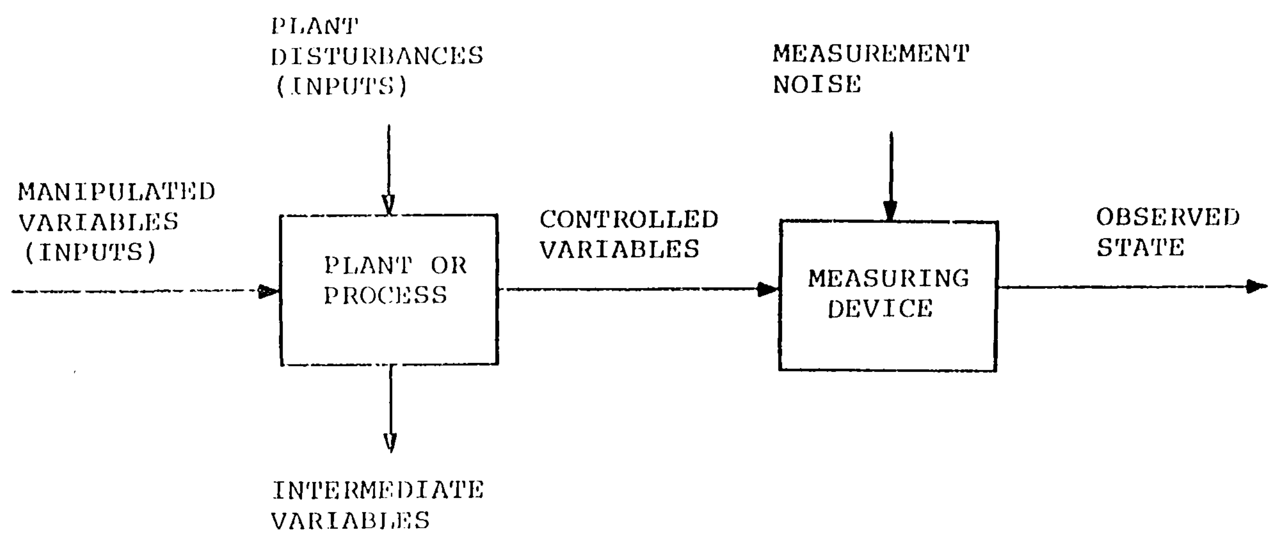

FIGURE 2.5 ELEMENTS OF A TYPICAL PROCESS

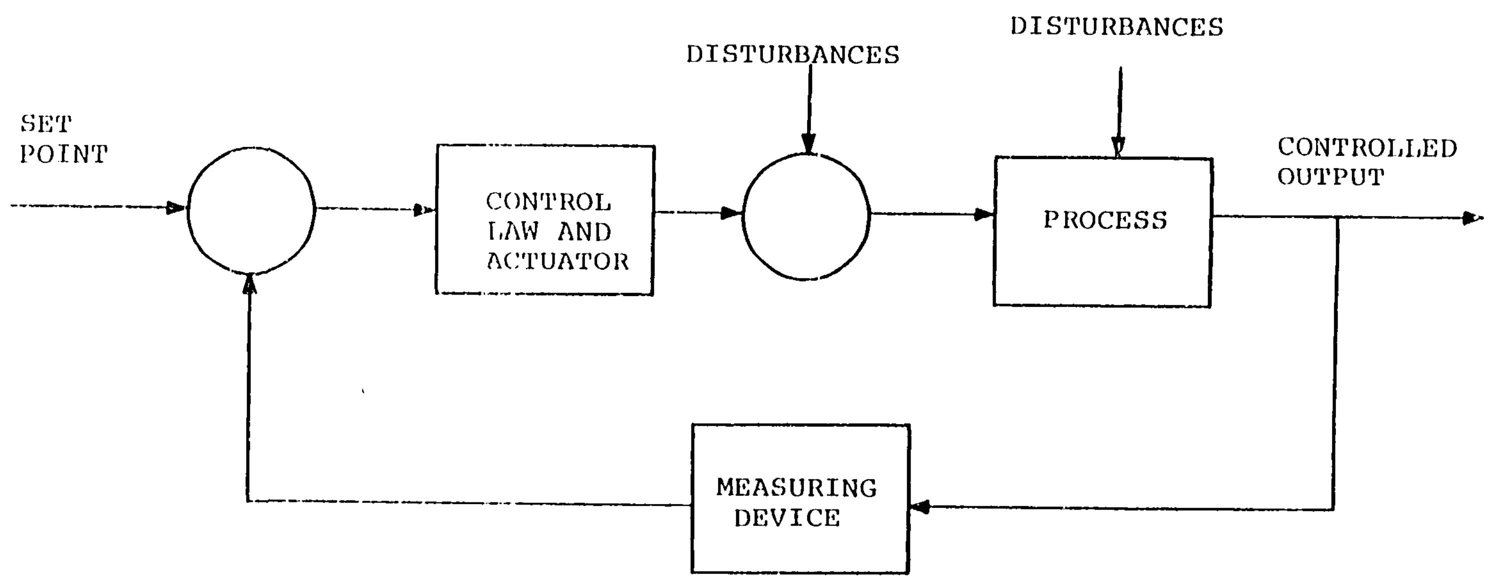

FIGURE 2.6 TYPICAL FEEDBACK CONTROLLER

Feedback controllers, however, have been widely used due to their simplicity and low cost.

Figure 2.7 is a diagram of a typical feedforward controller. In this case, the disturbances are measured (or predicted) and a model of the process (or process history) is used to provide the appropriate control actions. Feedforward controllers can theoretically provide perfect control if the disturbances are sensed perfectly and the process model is perfect. In practice this perfection is never achieved and feedforward controllers are often augmented with feedback as shown in Figure 2.8.

Control of the Activated Sludge Process

The aeration basin and the settler are two tightly coupled units. Therefore it is impossible to discuss the control of one unit without discussing its effect on the other. Also, it is necessary to delineate the possible control variables (manipulated variables), their control authority, and the time frame in which each works.

Influent Flow Rate Control

The influent flow rate affects the operation of the activated sludge process in many ways. Detention time in the aeration basins and the overflow rate of the settlers are direct functions of the influent flow rate. Additionally, the organic concentration of the influent generally varies almost in phase with the flow (79). This leads to an organic loading variation even greater than that of flow.

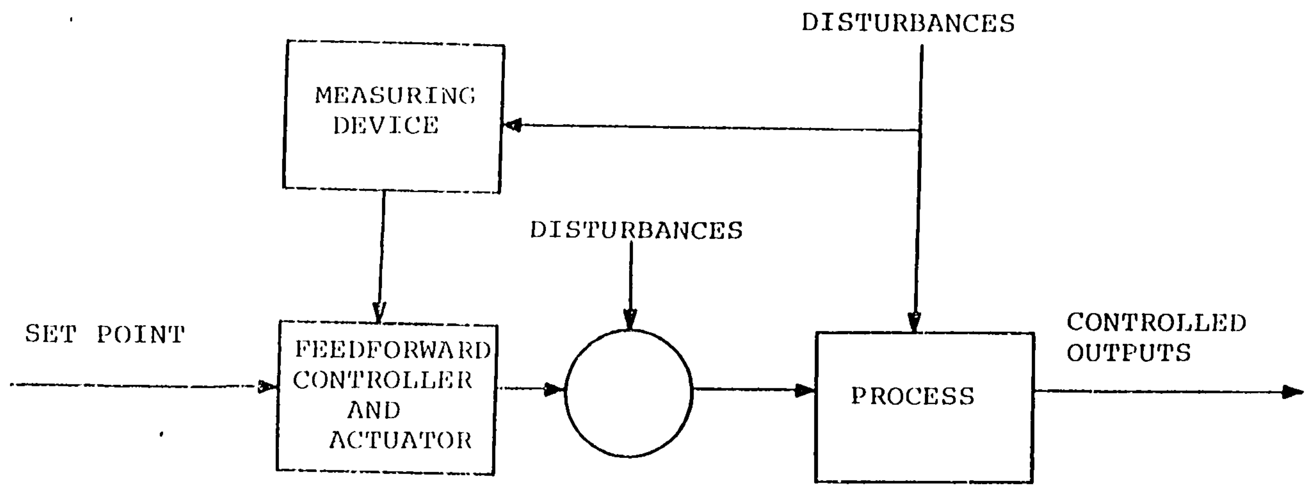

FIGURE 2.7 TYPICAL FEEDFORWARD CONTROLLER

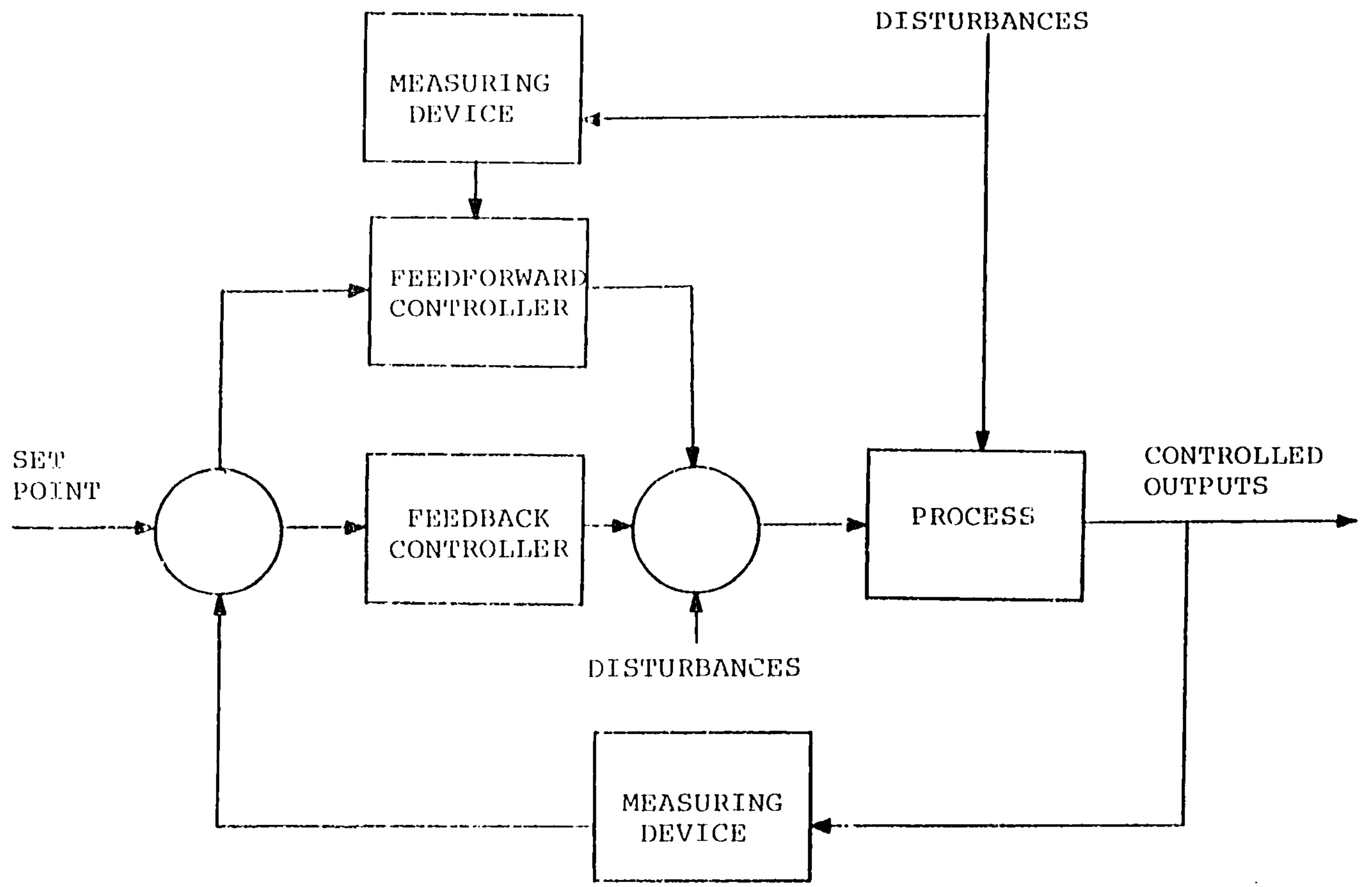

FIGURE 2.8 FEEDFORWARD CONTROLLER WITH FEEDBACK TRIM

Previous investigations have shown that wastewater treatment plants have little dampening capacity for hydraulic transients (14) and that these disturbances have immediate effects which are slow to die out (31). Olsson (87) concluded that there is no effective control action to minimize the effects of hydraulic shocks other than elimination of the shocks.

The influent flow rate is often not considered a control variable but a disturbance. This is because in most wastewater treatment plants all the flow that arrives must be treated. However, by using different control strategies for the influent pumping, it is possible to modify the shape of the flow curve and the rate of change.

Most control work concerning the influent flow rate has centered around flow and concentration equalization. Several papers (40, 84, 133) and an EPA design manual (125) have been published on design techniques for sizing these basins. Dold (41) developed an operational strategy for equalization which minimized both flow and load variations. Andrews, et al. (2) simulated the benefits of both constant volume and variable volume equalization on an autocatalytic reaction in a CSTR reactor and compared the results with those from various control strategies. Several papers (73, 104) have documented the benefits of equalization on primary and secondary settling. However, there is little evidence to support the hypothesis that flow equalization improves the biological aspects of wastewater treatment.

A variation of the influent flow equalization scheme has been the use of variable volume reactors with pumping to the settlers at a constant rate. Speece and LaGrega (104) developed this idea in comparison to other equalization alternatives. Poduske (96) stated that this feature was a key control tool in meeting effluent standards at the Tennessee Eastman plants.

Waste Sludge Flow Control

The waste sludge flow largely controls the total amount of solids in the activated sludge system. Olsson (86) has shown that the time constant for sludge wastage is approximately the sludge age. Thus, sludge wastage is a slow control variable with a time constant on the order of several days.

The waste sludge flow rate is usually controlled to maintain a constant F/M (food to microorganism) ratio or a constant sludge age. Both methods have been demonstrated to produce sludges with good removal efficiencies and settling characteristics. In steady state it can be shown that the sludge age and F/M ratio are approximately related (111).

$$1/SRT = Y\,(F/M) - Kd \qquad [2.25]$$

where:

SRT = sludge age (T),

Y = mass of solids produced per unit mass substrate removed,

Kd = specific organism decay rate (1/T).

Although F/M control is usually thought of as a steady state concept, an extension of this to the time domain is known as instantaneous F/M control. With this method, it is attempted to keep the ratio of solids under aeration proportional to the instantaneous organic loading. Flanagan (48) demonstrated that this objective was impossible without off-line storage of sludge. Using an off-line basin for storage of concentrated sludge, he derived control laws for recycle and waste flow rates to maintain a constant instantaneous F/M ratio. Cahion, et al. (19) derived similar control laws and demonstrated the technique at pilot-scale. They found that instantaneous F/M control did decrease the soluble organic material in the effluent but increased the suspended solids. The result was no net benefit.

By definition the sludge age is defined as the total mass of solids in the system divided by the mass of solids wasted in a given time period and usually has the units of days.

$$SRT = \frac{X_M}{Fe \cdot Xe + Fw \cdot Xw} \qquad [2.26]$$

where:

X_M = mass of solids in system (M),

Fe = effluent flow rate (L^3/T),

Xe = effluent SS concentration (M/L^3),

Fw = waste sludge flow rate (L^3/T),

Xw = waste sludge SS concentration (M/L^3).

Although Equation [2.26] can be easily solved for the waste flow rate, the suspended solids concentrations are usually not immediately available unless on-line instrumentation is used. Denniston, et al. (34) and Lukasik, et al. (75) used laboratory data with a three day time lag and developed a forecasting system utilizing exponential filtering which gave good results for predicting sludge wastage flows. In earlier work, Garrett (50) avoided the necessity of solids measurements by wasting a constant fraction of the aeration basin volume directly from the reactors.

Other sludge wasting control strategies have been developed. One demonstrated by Sorensen (105) is particularly simple and deserves further comment. In his pilot plant, the sludge blanket level in the clarifier was measured automatically. When the sludge blanket exceeded a certain level, the excess sludge pump would run for a fixed length of time. This strategy allowed Sorensen to operate his plant with a high solids concentration without the risk of losing the blanket over the weirs.

Return Sludge Flow Rate Control

The return sludge or recycle flow rate contributes to the solids loading to the clarifier and partially determines the underflow solids concentration from the settler. To a lesser extent the recycle flow rate also effects the distribution of solids between the settler and aeration basins. For underloaded clarifiers, increases in recycle flow rate result in decreases in underflow concentration and fewer solids stored in the settler. Decreasing the recycle flow has the opposite effect. Return sludge flow rate control is generally considered to have limited control authority with a time constant of hours.

There have been several theories on how to manipulate the recycle flow rate in the activated sludge process. For instance, Keinath, et al. (70) presented a scheme based upon a solids flux technique and a state point concept to minimize recycle pumping and maintain the settler at a critically loaded level. Flanagan (48) and Cashion, et al. (19) derived control laws for recycle flow to maintain the instantaneous F/M ratio as a constant. Wells, et al. (136) derived control laws suitable for manual implementation on a hand calculator which adjusted recycle flow to maintain a constant specific oxygen utilization rate (SCOUR). Stenstrom (110) also derived a control law for recycle flow in a multi-reactor configuration for minimizing the variability of SCOUR.

Despite the apparent advantages of the control strategies mentioned above, two control strategies are used almost universally. These strategies are the constant recycle flow rate and the flow proportioned recycle flow rate. This fact probably results from the lack of demonstrated, practical improvements due to other algorithms. With a constant recycle flow, both MLSS and RAS concentrations vary with influent flow. With flow proportioned recycle flow, both the MLSS and RAS concentrations remain relatively constant (70).

Feed Forward Flow Rate Control

Feed forward flow capability is used to change the contacting pattern in the activated sludge process by introducing raw (or settled) sewage at two or more points. This capability is often called step feed. This process modification was developed by Gould and first applied in New York City in 1939 (79). The use of step feed results in distributing the oxygen demand more uniformly throughout the aeration basins and also affects the distribution of solids in the different aeration basins.

Andrews and Lee (3) have shown by simulation that by introducing all the influent flow into the last aeration basin, the solids concentration going to the settler can be rapidly changed. Thus step feed has considerable control authority in affecting the solids distribution both between settler and reactors and between the different reactors. Step feed has a time constant of minutes to hours.

Although the step feed modification of the activated sludge process has been is use since 1939, much less has been written about its use in process control. One of the first documented uses of step feed (or step aeration as it was called at the time) was that of Torpey (120). Torpey's plant had a capacity of 40 MGD (151 M l/day) with the capability to feed settled sewage into any of four passes. When the settling characteristics of the sludge deteriorated, Torpey moved the feed point(s) toward the end of the process. This action increased the solids concentration in the first reactors, decreased it in the last reactors, and prevented solids from being lost over the weirs due to a high sludge blanket. As the settling characteristics improved, he moved the feed point(s) back toward the front of the plant. Torpey's method proved to be very effective in controlling sludge settling characteristics and preventing gross process failure. Andrews and Lee (3) predicted Torpey's findings by simulation and explored other contacting patterns.

Stenstrom (110) and Busby (16) have shown that SCOUR is directly related to growth rate (and therefore also to effluent quality). Stenstrom derived control laws for recycle flow and the step feed flow split in a multi-reactor system to minimize variations of SCOUR. Andrews, et al. (2) demonstrated this use of SCOUR as a controlled variable. Yust and Murphey (143) utilized Stenstrom's model and used an optimization technique to control the flow split in a three reactor pilot plant to minimize variations of SCOUR in the third reactor. They observed reduced effluent variability for soluble organic carbon but an increase in effluent suspended solids concentration. They contributed this solids increase to possible over aeration in the pilot plant due to low oxygen transfer efficiencies. This phenomenon will be discussed in a later section.

Andrews, et al. (4), in pilot plant work with the City of Houston, demonstrated the use of step feed for effluent ammonia control. By regulating the effluent ammonia concentration, it was possible to avoid breakpoint chlorination. The resulting chlorine savings anticipated at the full-scale plant was estimated to be $250,000 annually. Sorensen (105) reported on the use of step feed to control effluent BOD in parallel pilot plants. He was able to decrease the effluent variability, save oxygen, and decrease sludge production with the use of step feed.

Air Flow Rate

The air flow rate controls the dissolved oxygen in the aeration basins. There is usually a lower limit on the air flow to maintain adequate mixing and prevent settling of solids in the basins. The upper limit may be determined by mechanical constraints or floc breakup (92, 93). Since the activated sludge process depends on aerobic biological reactions, the dissolved oxygen concentration affects the whole process. High DO concentrations promote fast reaction rates (89) but result in higher operational costs since the mass transfer of oxygen is a function of the oxygen deficit. Low DO concentrations have been demonstrated to have deleterious effects on the effluent suspended solids (108) and nitrification (82). There is still much controversy in the field over the DO level necessary for proper operation. Several texts (79, 90) list 1.0-2.0 mg/l as a reasonable level, but other researchers (67) have suggested lower concentrations. The aeration system has a time constant of 15 to 30 minutes in most treatment plants.

Aeration control is presently practiced in many wastewater treatment facilities. Since aeration costs often amount to 50 to 80 percent of the entire power consumption at activated sludge plants, considerable effort has been exerted to establish air flow rate control strategies for minimizing power consumption. Most of these strategies utilize feedback controllers with DO being the measured variable. EPA published a design manual (126) on DO control in 1977. Aeration control is not the subject of this investigation and will not be discussed further except as it may influence the concentration of effluent suspended solids.

III. EXPERIMENTAL PROCEDURES AND EQUIPMENT

Description of the Sagemont Wastewater Treatment Facility

The research reported in this dissertation was performed at the City of Houston's Sagemont wastewater treatment facility. The plant treats a waste almost entirely of domestic origin with only a small commercial and no industrial contribution. It is designed for a dry weather flow of 5 MGD (18.9 M l/day) and has a maximum hydraulic capacity of 25 MGD (94.6 M l/day). The current effluent permit requirements are shown in Table 3.1.

In 1981 the plant had an average flow of 2.3 MGD (8.7 M l/day) and discharged an effluent with yearly average suspended solids, BOD_5, and ammonia nitrogen concentrations of 6, 3, and 0.2 mg/l, respectively. Peak flows were in excess of 20 MGD (75.7 M l/day) and occurred during severe rain storms.

A plan view of the Sagemont plant is shown as Figure 3.1. The majority of the influent flow enters through the pumping station located on-site. After screening, the raw wastewater is introduced at the head of the aeration basins where it is mixed with return activated sludge. The MLSS flows through the of aeration basins and is settled in sedimentation tanks. The thickened underflow is returned to the head of the plant while the overflow is chlorinated before discharge. Waste activated sludge is pumped to a thickener and aerobic digesters for further treatment.

TABLE 3.1

PERMIT REQUIREMENTS FOR THE SAGEMONT WWTF

Item	Maximum 7-day Average	Maximum 30-day Average
Suspended Solids (mg/l)	20	12
BOD_5 (mg/l)	10	5
Ammonia Nitrogen (mg/l as N)	10	2

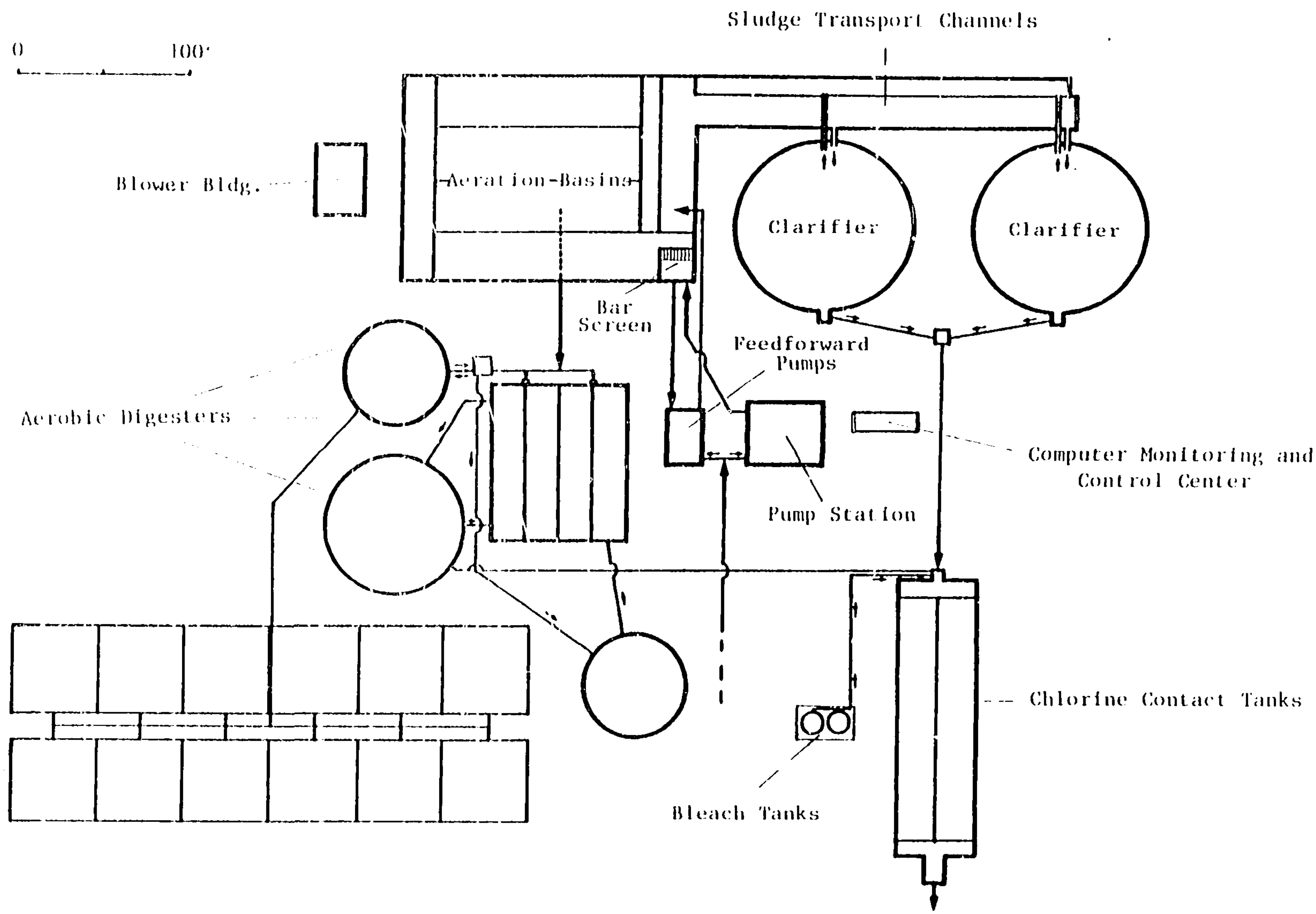

FIGURE 3.1 PLAN VIEW OF THE SAGEMONT WWTF

Two major plant modifications were necessary to conduct this research. The first was to take two aeration basins, one clarifier, and one chlorine contact basin out of service. This was necessary to increase the plant loadings to near design capacity. Electric actuators were installed on the sluice gates to the clarifier and chlorine contact basins so that they could be brought back on-stream during times of high flow.

The second modification was the repiping of an old influent pump station so that a portion of the influent wastewater could be diverted to the third aeration basin. This is described in more detail later.

Pump Station

The pump station has three pumps, each equipped with a different type of control system. These control systems, all of which are based on liquid level, are: 1) variable speed pumping via a variable frequency drive, 2) automatic throttling of the pump discharge, and 3) on-off, fixed speed pumping. An interface to the pump controller allows the process control computer to override the built-in control algorithms and take direct control of the pumps for experimental purposes. This capability was used to generate specific flow forcings.

The pump station is a two-level, wet-well/dry-well design. The motors and motor control center are located on the upper level, and the centrifugal pumps are located on the lower level. The volume of the wet well between high and low levels is approximately 38,000 gallons (144,000 l), giving a detention time of 11 minutes at design flow. Pump capacities are 4000 gpm (21.8 M l/day) for the variable flow pumps and 7000 gpm (38.2 M l/day) for the on-off pump.

Feed Forward Pumps

The piping of an old pump station, located at the plant, was modified to give partial feed forward capability. The purpose of the feed forward pumps is to introduce the raw wastewater at a point other than the first pass of the aeration basins. This capability can be used to modify the DO and suspended solids concentration profile through the aeration basins.

As shown in Figure 3.1, the feed forward pumps draw raw wastewater from the bar screen area and pump it to aeration basin 3. Each pump is an 8-inch self-priming design with a capacity of 1200 GPM (6.5 M l/day). A motor operated pinch valve was installed on one pump to modulate the flow.

Aeration Basins

Prior to the commencement of this study, the Sagemont activated sludge process was operated in an aeration-reaeration mode with both volumes being approximately equal. For the purposes of this study, the plant was reconfigured by taking two basins out of service and routing the flows as shown in Figure 3.2. This is the conventional activated sludge process. This modification was made to increase the volumetric loading of the aeration basins to near design loading. The approximate dimensions and volumes of the aeration basins are given in Table 3.2

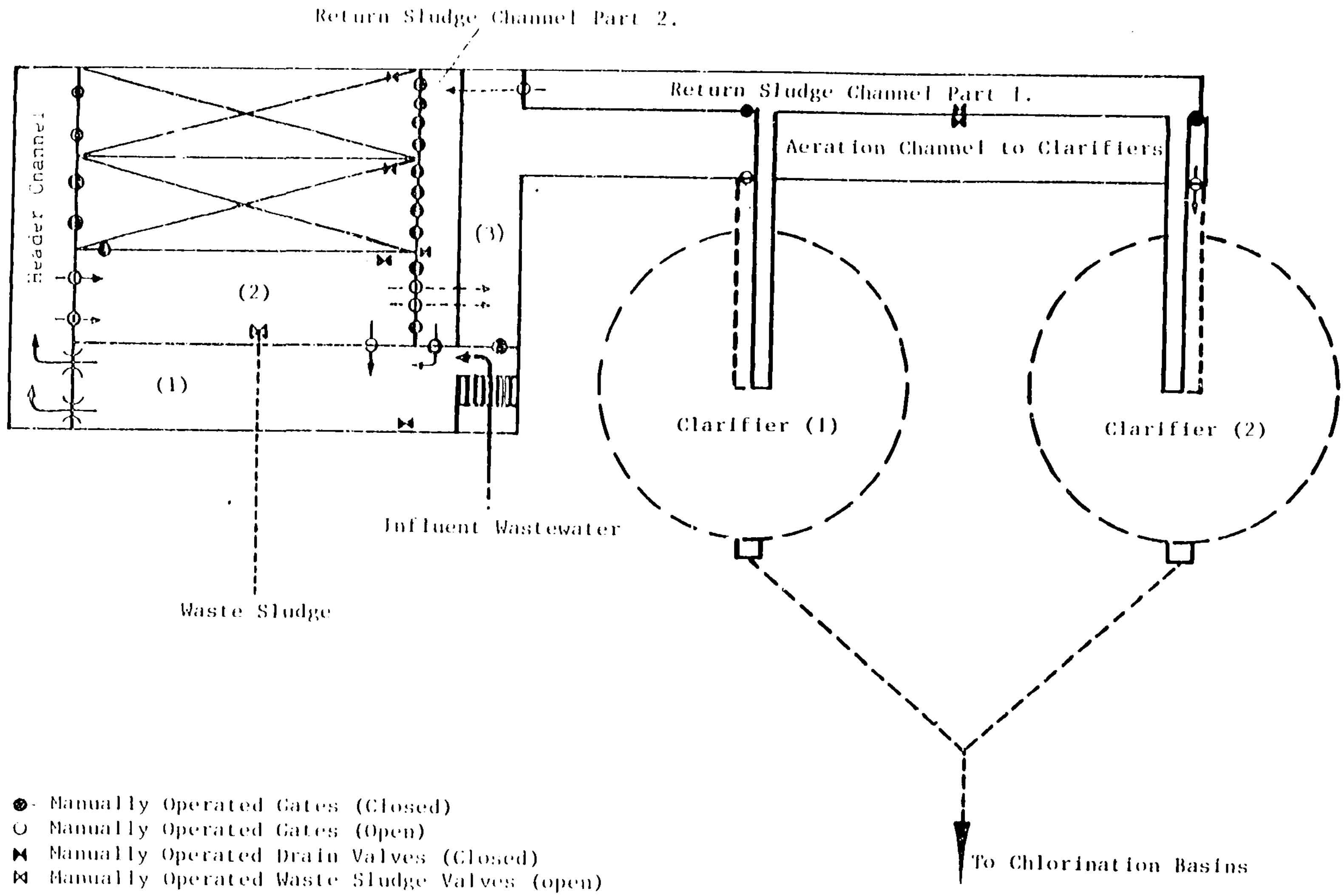

FIGURE 3.2 CONFIGURATION OF AERATION BASINS DURING THIS STUDY

TABLE 3.2

DIMENSIONS AND VOLUMES OF AERATION BASINS

Basin	Dimensions Ft	Volume MG (Cu M)
Aeration Basin 1	30 x 122.5 x 14.5	0.41 (1570)
Header Channel	20 x 123.5 x 14.5	0.27 (1014)
Aeration Basin 2	30 x 110 x 14.5	0.36 (1375)
Aeration Basin 3	20 x 95 x 14.5	0.21 (780)
Aeration Basin 4	20 x 223 x 8	0.13 (505)[1]
RAS Channel 1	10 x 213 x 8	0.13 (480)
RAS Channel 2	10 x 95 x 14.5	0.10 (390)

[1] Based upon 1/2 true volume due to grit deposits.

Four centrifugal blowers provide compressed air for the aeration basins, the airlift pumps used for sludge recycle, and mixing in the chlorine contact basins. Each blower is rated for 3900 SCFM at 7.0 psi discharge pressure. Two of the blowers are equipped with motor operated valve actuators on the suction valve to modulate air flow rate. The process control computer also has on-off control of all four blowers.

Waste sludge can be withdrawn from either aeration basin 1 or 2 as shown in Figure 3.2. Manual valves are provided to select the basin from which sludge is to be withdrawn. Withdrawal is on an intermittent basis (approximately four hours per day at a sludge age of ten days) using two centrifugal pumps to feed the excess sludge to a thickener or directly to the aerobic digesters.

Clarifiers

The two clarifiers at Sagemont are each 100 feet (30 m) in diameter with 18 foot (5.5 m) diameter inlet wells which extend 6 feet (2.8 m) below the water surface. The nominal side water depth is 12 feet (3.7 m) at a flow rate of 10 MGD (38 M l/day). The effluent is removed by 570 linear feet (175 m) of double-sided V-notch weir. A plan view of one of the clarifier showing interconnections to the aeration basins is shown in Figure 3.3. Cross sections giving the details of the inlet and outlet structures are shown in Figures 3.4 and 3.5.

One of the two clarifiers was been taken out of service for the duration of this study. This doubles the loading on the on-stream clarifier.

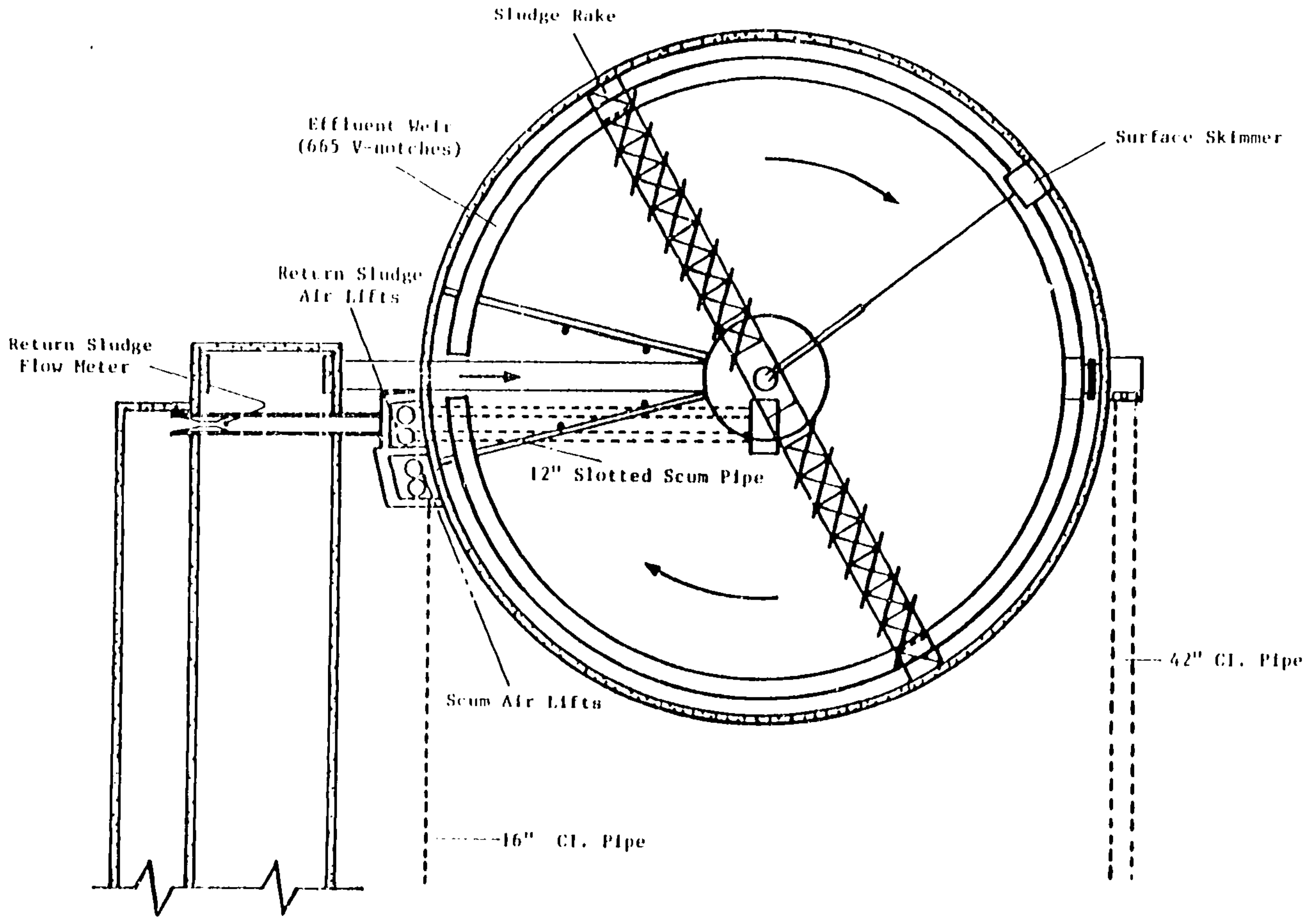

FIGURE 3.3 PLAN VIEW OF CLARIFIER

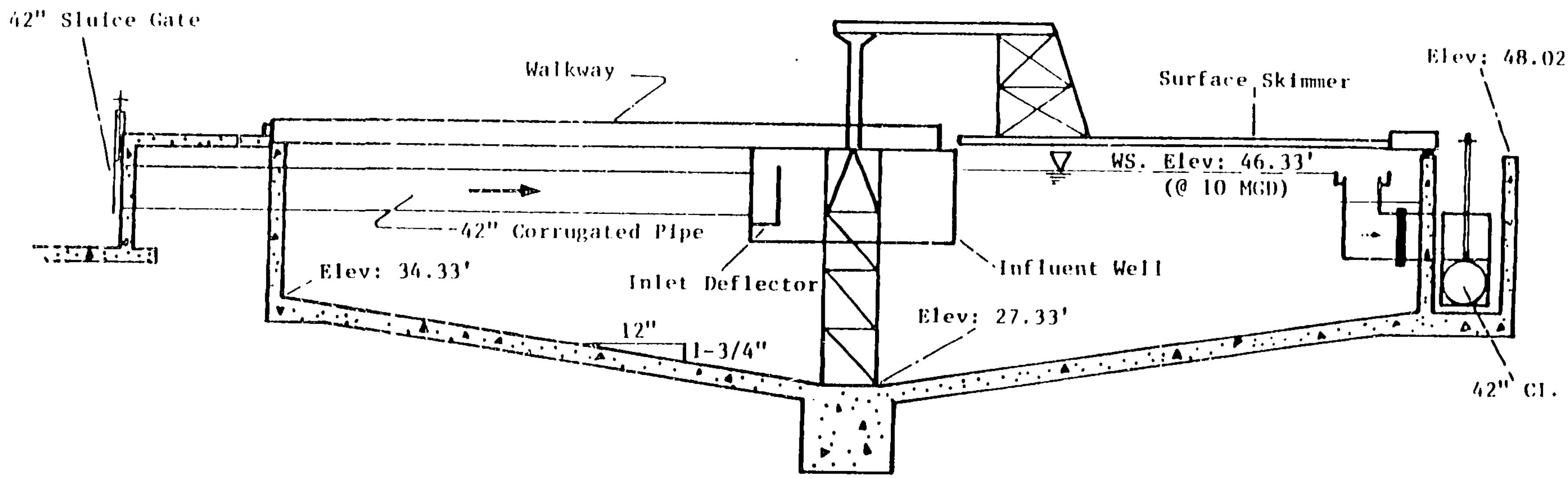

FIGURE 3.4 CLARIFIER INLET AND OUTLET PIPING

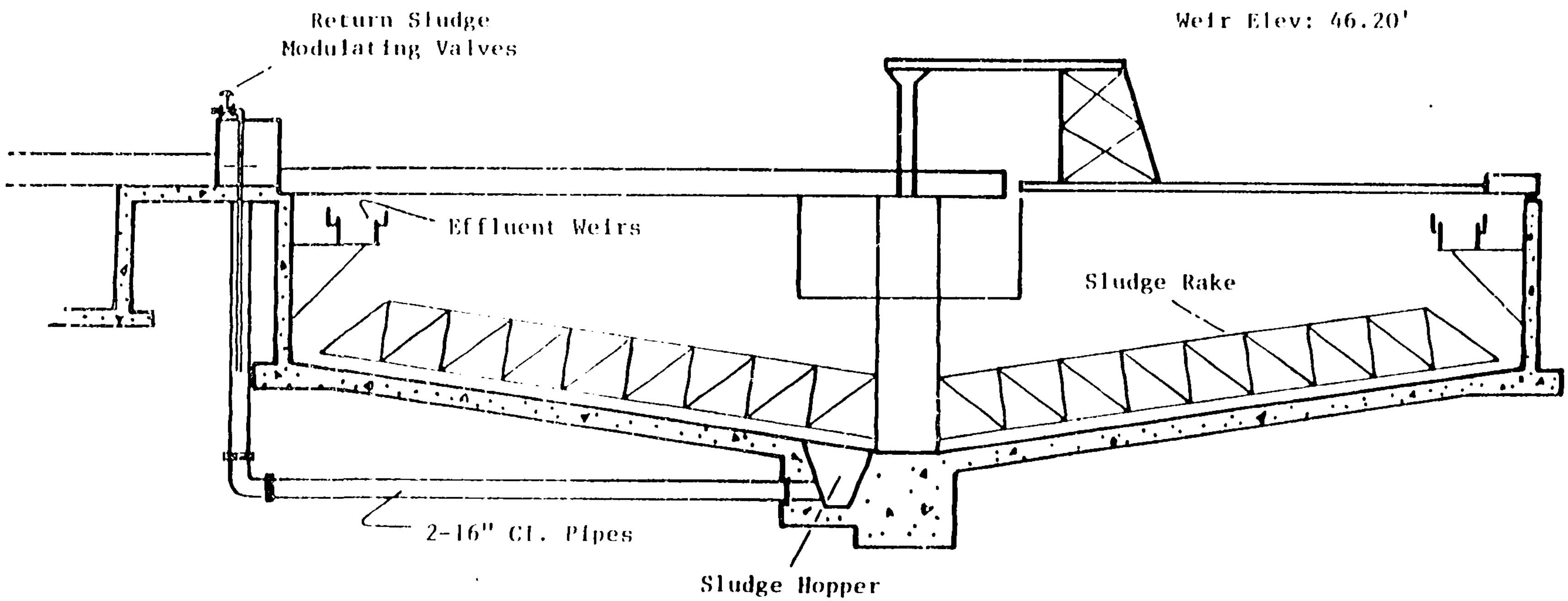

FIGURE 3.5 CLARIFIER WEIRS AND SLUDGE REMOVAL MECHANISMS

The maximum design flow for each clarifier is 12.5 MGD (47 M l/day) giving an overflow rate of 1,600 gpd/sq ft (2.7 m/hr). Since the flow at Sagemont is frequently greater than this, provision has been made to automatically bring the other clarifier back on-stream whenever the plant flow exceeds 12.5 MGD.

The MLSS enters the settler through a sluice gate leading to a 42-inch (1.07 m) corregated pipe as shown in Figure 3.4. The sludge strikes a deflector plate in the center well which diverts it around the well. The clarifier has a sloped bottom (1-3/4 to 12) and rake to move the thickened sludge to an off-center sludge hopper. Sludge is removed from the hopper by two 16-inch (0.41 m) airlift pumps. Each airlift has a 4-inch (10 cm) air line controlled by a butterfly valve. Electric actuators have been installed on the air valves for the instrumented clarifier to permit computer control of the return sludge flow rate over a range of 0 to 6 MGD (0 to 22.7 M l/day). The airlifts discharge into a 30 foot (9.1 m) long, 24" x 30" channel. A 9-inch (22.9 cm) Parshall flume with free discharge to the return sludge channel is installed for flow measurement.

The scum removal system is shown in Figure 3.3. It consists of a reciprocating surface skimmer, two adjustable 12-inch (30.5 cm) slotted pipes, and two 6-inch (15.2 cm) airlift pumps. The surface skimmer sweeps the surface of the settler and pushes the floating material to the slotted pipes. The airlifts pumps the scum from the slotted pipes to a scum box. From there, flow is by gravity to a manhole and eventually back to the influent pump station.

Computer Monitoring and Control System

During the course of this research, a complete computer monitoring and control system was designed, installed, and tested at the Sagemont WWTF. The system consists of a process control computer and its peripherals, three programmable controllers, and a table-driven data acquisition and control package. Since the system includes both a central computer and intelligent field controllers, it can be classified as a distributed control system.

Process Control Computer

The process control computer selected for the project was a Digital Equipment Corporation (DEC) PDP-11/23. A schematic of the computer hardware and its peripheral equipment is shown in Figure 3.6. A list of computer hardware and peripherals is presented in Table 3.3.

System Software

The operating system selected for the project was RSX-11M. It is a multi-user, multi-tasking, real-time operating system which is well adapted for process monitoring and control. Two computer languages were provided on the Sagemont computer, MACRO-11 and FORTRAN-77. The FORTRAN compiler produced direct PDP-11 machine code optimized for execution-time efficiency on a PDP-11 with a floating point processor.

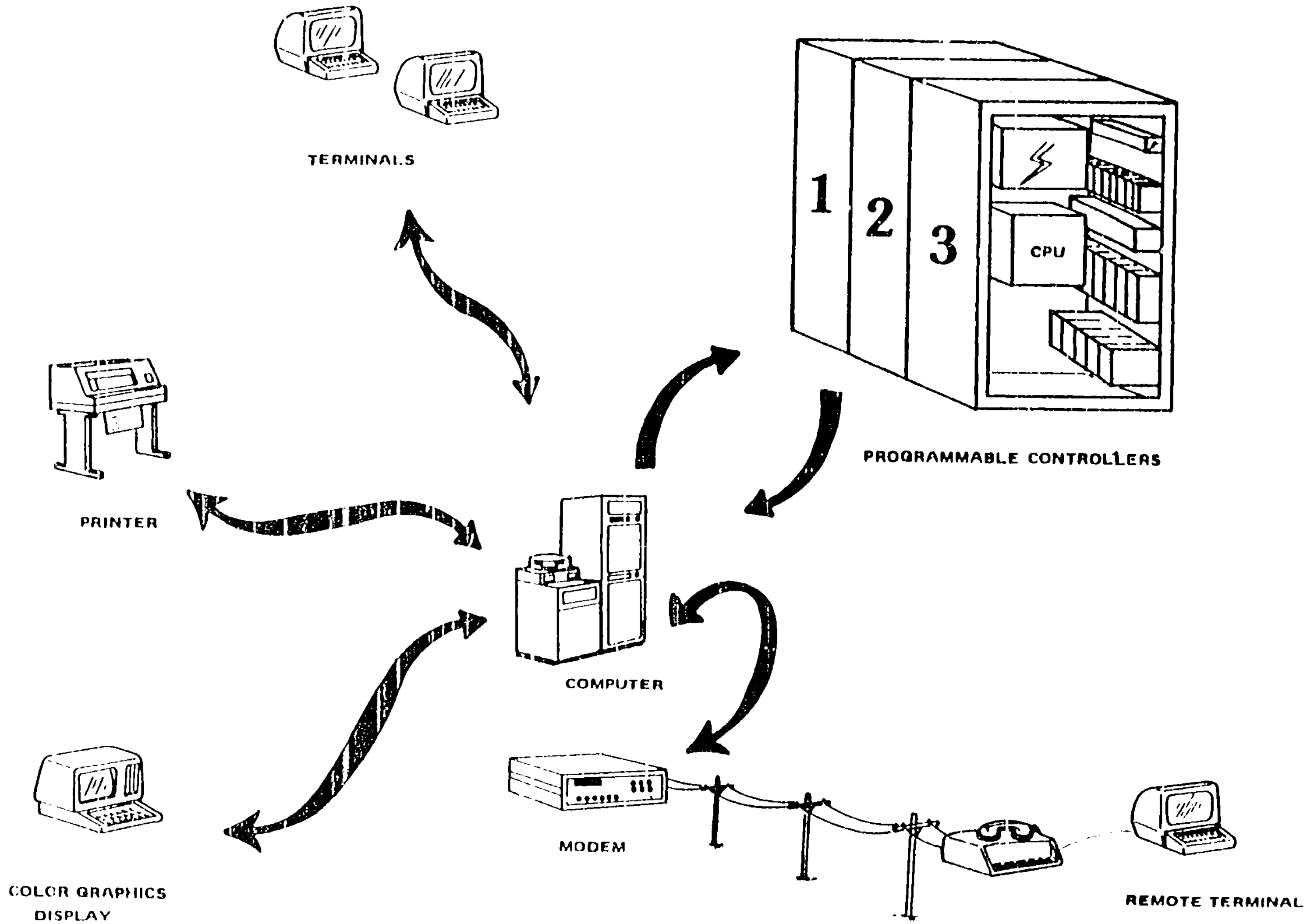

FIGURE 3.6 COMPUTER HARDWARE SCHEMATIC

TABLE 3.3

SAGEMONT COMPUTER HARDWARE AND PERIPHERALS

* LSI-11/23 CPU with Memory Management Unit and Floating Point Option
* Four MSV11-DD 64 Kb RAM cards
* Dual RL02 10.4 Mb cartridge disks with RLV11 controller
* Dual RX02 512 Kb floppy disks with RXV21 controller
* TSV05 tape transport and controller
* Three DLV11-J four serial line interfaces
* DZV11-B four serial line asynchronous multiplexer
* DRV11-J high density parallel line interface
* KWV11-A programmable crystal clock
* IBV11-A IEEE-488 instrument bus interface
* BDV11-A Boot/Diagnostic/PROM,ROM/Terminator card
* DEC VT125 CRT Graphics Terminal
* Two Micro-Term MIME-2A CRT Terminals
* DEC LA120 Printing Terminal
* Epson MX-80 Printer
* Racal-Vadic VA255 Auto-Answer Modem
* Two ISC 8951 Color Graphics Computers

Programmable Controllers

Since the PDP-11/23 process control computer itself was not well adapted for interfacing with process signals, three Texas Instruments PM550 (117) programmable controllers (PC) were purchased for this purpose. The PM550 is an industrial grade PC based on the use of dual microprocessors.

Programmable Controller Interface

A proprietary software interface package was purchased from Texas Instruments to provide networking capability between the PDP-11/23 and the PM550's. The software driver is written in MACRO-11 assembly language while most other subroutines are written in FORTRAN IV. Communications between the PDP-11/23 and each PM550 was on a point-to-point basis over a dedicated four-wire cable.

Data Acquisition and Control Software

An overview of the data acquisition and control system developed for the Sagemont plant is shown in Figure 3.7. This figure shows the overall structure of the package, many of the data structures, and relationships between the various programs. As a whole, the on-line data bases and programs that act upon them formed a complete table-driven data acquisition and control system.

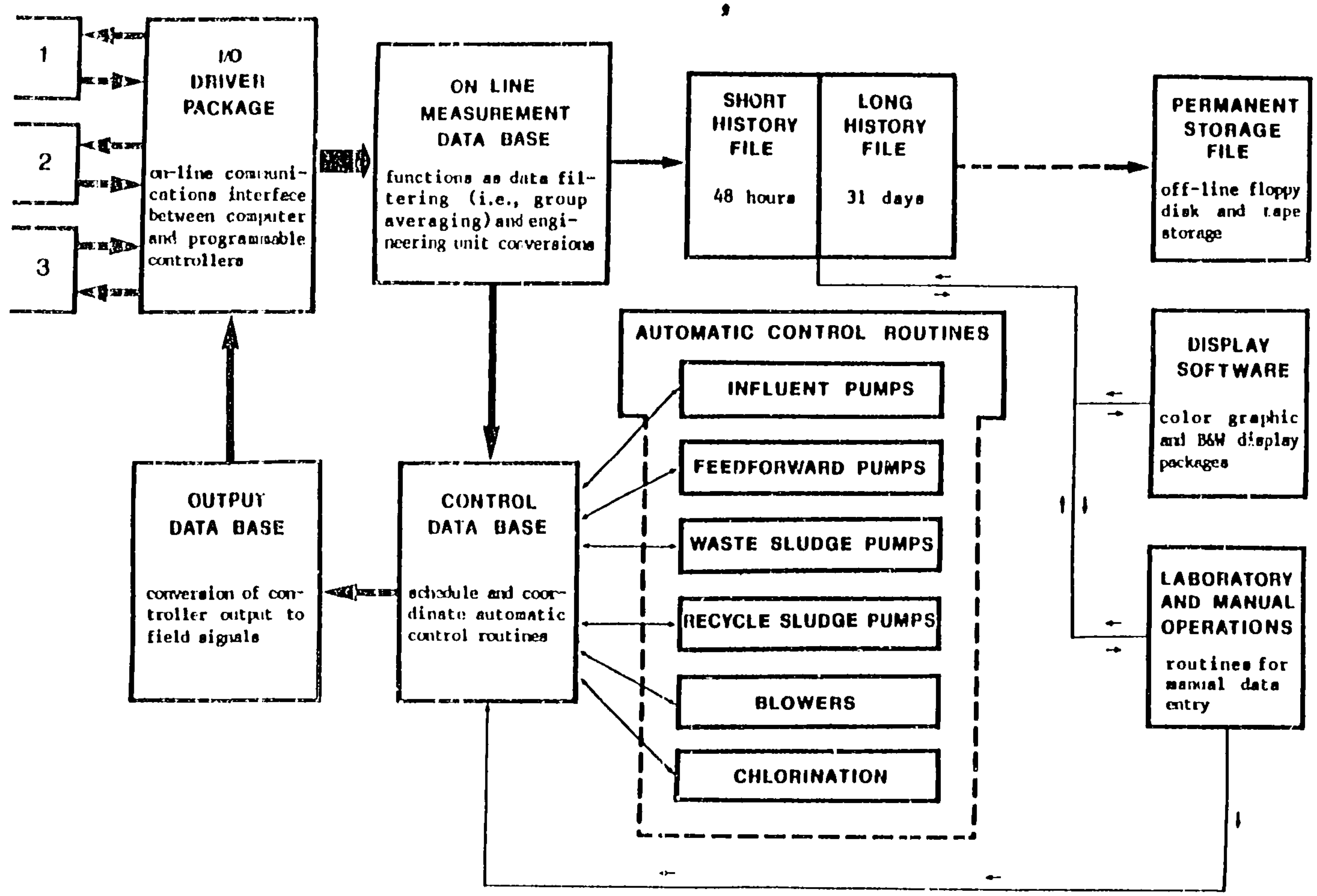

FIGURE 3.7 COMPUTER SOFTWARE SCHEMATIC

On-Line Data Bases

Data which was accessed frequently was retained in computer memory for rapid access by all the programs that read or modify the data. These data were organized into three data bases, a raw data area, and a header area. The raw data area was updated every six seconds.

The measurement data base can be thought of as 100 entries in a table, one for each measurement. Each entry in the table includes several flags, equation types and coefficients for filtering and conversion to engineering units, and limits for alarms. Most measurements were one minute averages.

The controller data base can be thought of as 20 entries in a table, one for each controller. Each entry includes several control flags, a task name, data and time entries, pointers to other data bases, and eleven controller specific variables. Controllers could be scheduled to run as frequently as once every 15 seconds and as little as once per 120 days.

The output data base includes the output value in engineering units, the type and address of the output, upper and lower limits of the output, and variables for converting the engineering units to field signals. These conversions can be used to implement pump and/or valve characteristics. Outputs can be a single bit (turn pump on or off), an integer, or a floating point number (setpoint to a PID control loop).

The data acquisition system logged one minute averages of continuous signals and one minute snap shots of discrete data onto a disk file for permanent off-line data storage.

On-Line Instrumentation

Like any computer monitoring and control project, the Sagemont project required extensive instrumentation. The system had 75 analog inputs, 8 analog outputs, 48 digital inputs, and 24 digital outputs. Only those instruments listed in Table 3.4, however, were directly utilized in this research and are described below. Figure 3.8 shows the location of many of these instruments.

Effluent Flow Meter

Effluent flow rate was measured with a V-notch weir located at the end of the chlorination basin. The water height is measured with a Wallace & Tiernan series 83-030 ball float. This instrument outputs a 4-20 mA signal proportional to the flow rate (0 to 16 MGD).

Return Sludge Flow Meter

The return sludge flow rate was measured through a 9-inch (22.9 cm) Parshall flume. The liquid depth in the flume is proportional to the back-pressure on a submerged tube through which air is bubbled. A purge regulator maintains a constant air flow. The flow rate, of course, is a well known function of liquid height. The instrument was fabricated using a copper bubble tube, a Fischer & Porter model 53RB2110 purge regulator with rotometer, and a Rosemount model 1151 DP transmitter. The DP transmitter outputs a 4-20 mA signal linear with level. The flow rate is calculated in the PDP-11/23 and has a range of 0 to 6 MGD (0 to 22.7 M l/day).

TABLE 3.4

ON-LINE INSTRUMENTATION

Measurement	Stream or Location
Flow	Effluent
Flow	Return Sludge
Flow	Feed Forward
Flow	Waste Sludge
Suspended Solids (3)	MLSS and RAS
Turbidity	Clarifier Effluent
Sludge Blanket Level	Settler
Temperature (3)	MLSS, RAS, Effluent
D. O. (4)	Aeration Basins 1, 2, 3, & 4

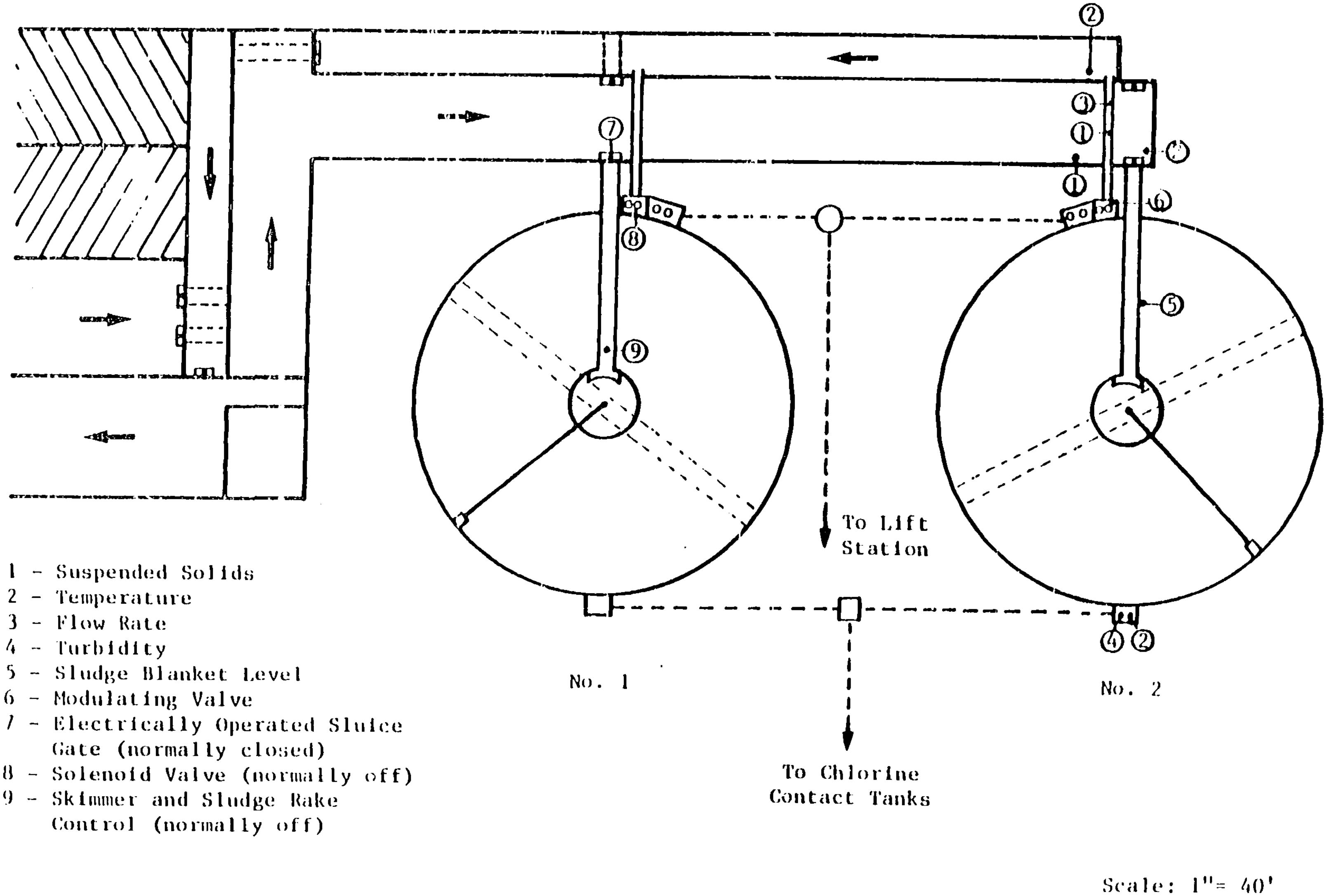

FIGURE 3.8 LOCATION OF ON-LINE INSTRUMENTATION

Feed Forward Flow Meter

The feed forward flow rate was measured in a 12-inch (30.5 cm) cast iron pipe with a Polysonics model DHT-HYDRA ultrasonic flow meter. This meter works on the Doppler principle and has a range of 0 to 5 MGD (0 to 19 M l/day). The meter output is a 4-20 mA signal proportional to flow.

Waste Sludge Flow Meter

The waste sludge flow rate was measured in a 12-inch (30.5 cm) PVC pipe with a Mapco model 1181 doppler flow meter. The output is a 4-20 mA signal corresponding to 0 to 600 gpm (0 to 2270 l/min).

Suspended Solids Concentration

Suspended solids were measured in aeration basin 2, aeration basin 4 leading to the clarifier, and the return activated sludge (RAS) channel. EUR-Control model MEX-2 meters with TAG 30/15 probes for MLSS and a TAG 10/5 probe for RAS were used. The meters work with infrared light on a 4-beam, alternating lamp principle which compensates for uneven coatings on the optical surfaces. The range is 0 to 10,000 mg/l for the MLSS probes and 0 to 20,000 mg/l for the RAS probe. The output is 4-20 mA proportional to suspended solids concentration. Figure 3.9 shows typical results of the meter compared to laboratory analysis.

Effluent Turbidity

The clarifier effluent turbidity was measured with a Sigrist Photometer model KTJ-25 coupled with a Great Lakes Instruments model 48 display

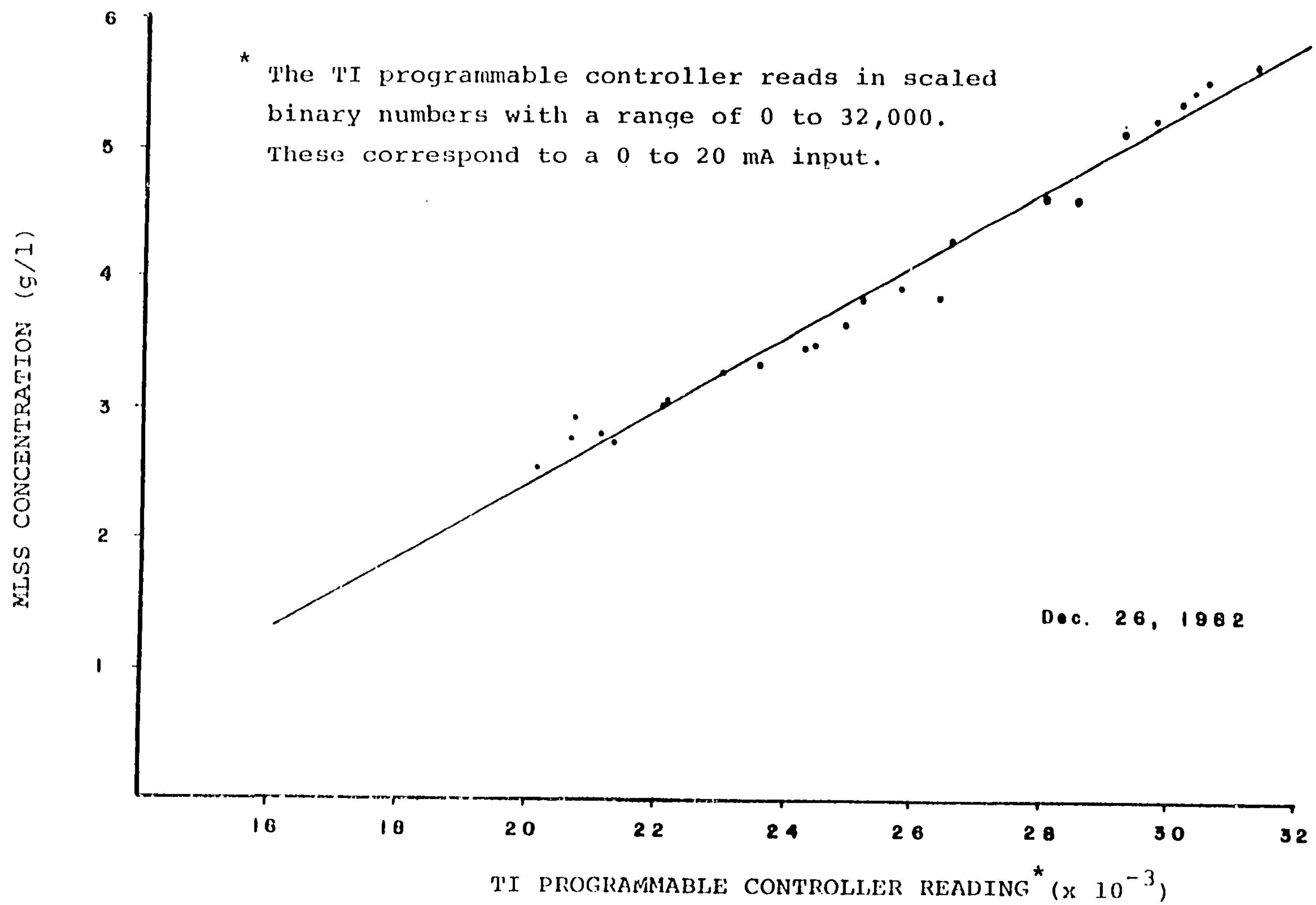

FIGURE 3.9 CALIBRATION CURVE FOR MLSS METER

unit. Turbidity can be correlated with effluent suspended solids as shown in Figure 3.10. Inside this complicated instrument, a lamp transmits a light beam to an oscillating (600 Hz) mirror which produces alternately a measuring beam and a reference beam. The reference beam passes through an internal turbidity standard (20 FTU), the sample stream, and reaches the photocell. The measuring beam passes through the sample, and light scattered at 25 degrees reaches the photocell. Thus, the photocell alternately receives two beams of different intensities. This difference in intensities drives a variable shutter which varies the intensity of the reference beam. At equilibrium, the shutter position is a measure of turbidity. The instrument transmits a 4-20 mA signal proportional to turbidity.

Sludge Blanket Level

The sludge blanket level and solids concentration profiles in the settler were measured with a EUR-Control model MEL-1 instrument with a TAG 10/5 probe. The MEL-1 works on the same principle as the MEX-2 solids meter except that a reel is used for automatically lowering and raising the probe. This instrument was modified for computer control to allow measurement of the solids profile while avoiding entanglement of the probe in the sludge rake. The instrument was mounted on the clarifier bridge approximately 25 feet (7.6 m) from the center. The instrument outputs two 4-20 ma signals proportional to probe depth (0 to 25 feet) and solids concentration (0 to 20,000 mg/l).

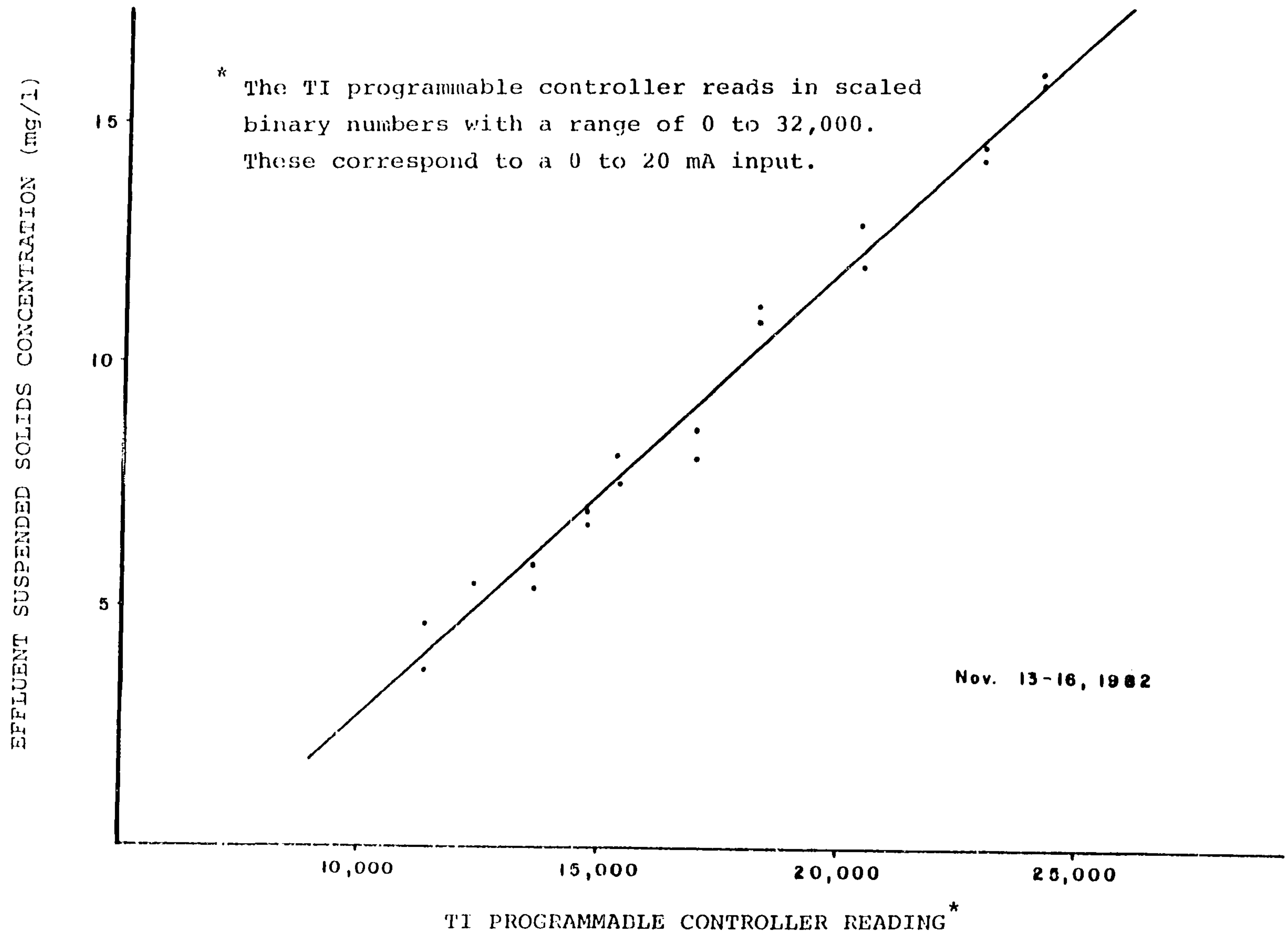

FIGURE 3.10 CALIBRATION CURVE FOR TURBIDIMETER

Temperature

Temperature was measured in aeration basin 4, the return sludge channel, and the clarifier effluent with Rosemount model 444 Alphaline RTD temperature transmitters with a 100 ohm RTD probe. The transmitters were all calibrated for a range of 0 to 40 degrees Celcius and have an 4-20 mA output proportional to temperature.

Dissolved Oxygen

Dissolved oxygen was measured in aeration basins 1, 2, 3, and 4 with four Leeds and Northrup model 7931 DO transmitters. The meters and probes were mounted on handrails approximately midway in the basins. The meters have a 4-20 mA output proportional to dissolved oxygen concentration.

Analytical Techniques

All analytical determinations were performed in compliance with the following procedures.

Tracer Tests

Rhodamine WT dye was used for the tracer tests reported in this research. This dye is reported (124) to degrade slowly and not readily adsorb onto activated sludge solids. Several tests confirmed this. However standards prepared with filtered wastewater and deionized water had a slightly different slope. This could be caused by color in the wastewater (124). Therefore, all standards used during the tracer tests were prepared with filtered wastewater. Since temperature greatly affected the results, all samples and standards were allowed to stand overnight for the temperatures to equilibrate.

Using a Perkin Elmer model 204 fluorescence spectrophotometer, it was possible to analyze for concentrations as low as 1 ppb. The maximum dye concentrations encountered in the tests were 50 to 70 ppb. Table 3.5 lists the instrument settings used in the analysis.

Suspended Solids

Suspended solids were analyzed in accordance with Standard Methods (107) with the exception that a microwave oven was used to dry the sample. Drying time was ten minutes. A rotating platform inside the oven was used to insure uniform heating of all the samples. Up to eight samples were dried at the same time. Each sample was dried and stored in an individual desiccator. All samples were run in duplicate. Due to the low suspended solids concentrations in the clarifier effluent, up to 2 liters per sample were used.

Settling Tests

All settling tests were performed in a three liter, stirred settling apparatus as described by White (137). This device is a 10 cm diameter plexiglas tube with a 50 cm settling height and a 1 RPM stirrer. Results from this test are claimed to be more indicative of true settling characteristics than the commonly used SVI test (138).

TABLE 3.5

Fluorescence Spectrophotometer Settings

Exciter Wavelength	550 nm
Analyzer Wavelength	575 nm
Sensitivity Control	10
Selector	X10
Light Source	Xenon Lamp

For the purposes of this research, the settling tests were used to determine the thickening characteristics of the activated sludge. A series of settling tests with sludges of different concentrations were performed. In each test the sludge height and time were recorded and plotted as in Figure 3.11. The slope of the linear portion of this plot is known as the initial settling velocity (ISV) and is assumed to be a function only of the solids concentration. The results of several series of these tests are reported in a later section.

The data from these tests can be fitted to an exponential equation with two unknown coefficients, V_0 and b.

$$Vs = V_0 \cdot \exp(b \cdot X) \qquad [3.1]$$

where:

V_0, b = settling parameters.

Equation [3.1] can be used directly in a mathematical model or used to derive a settling flux curve (Figure 3.12) for design purposes.

Parameter Estimation

A direct least squares technique and a Simplex optimization technique were used to estimate model parameters in this investigation.

Direct Least Squares

Whenever it was possible to express the equations in the form of Equation [3.2], a direct least squares solution (Equation [3.3]) was used.

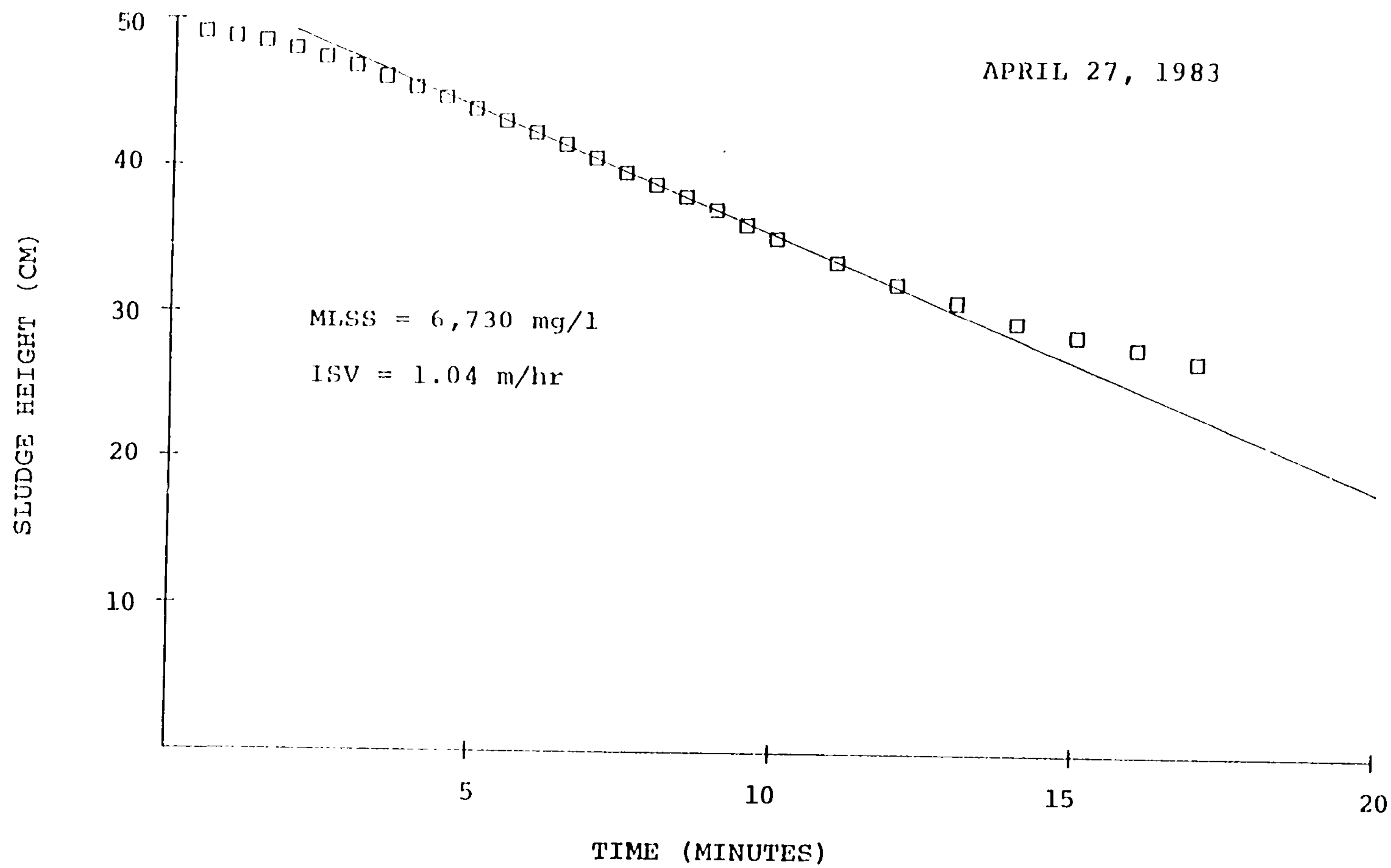

FIGURE 3.11 EXAMPLE SLUDGE SETTLING TEST

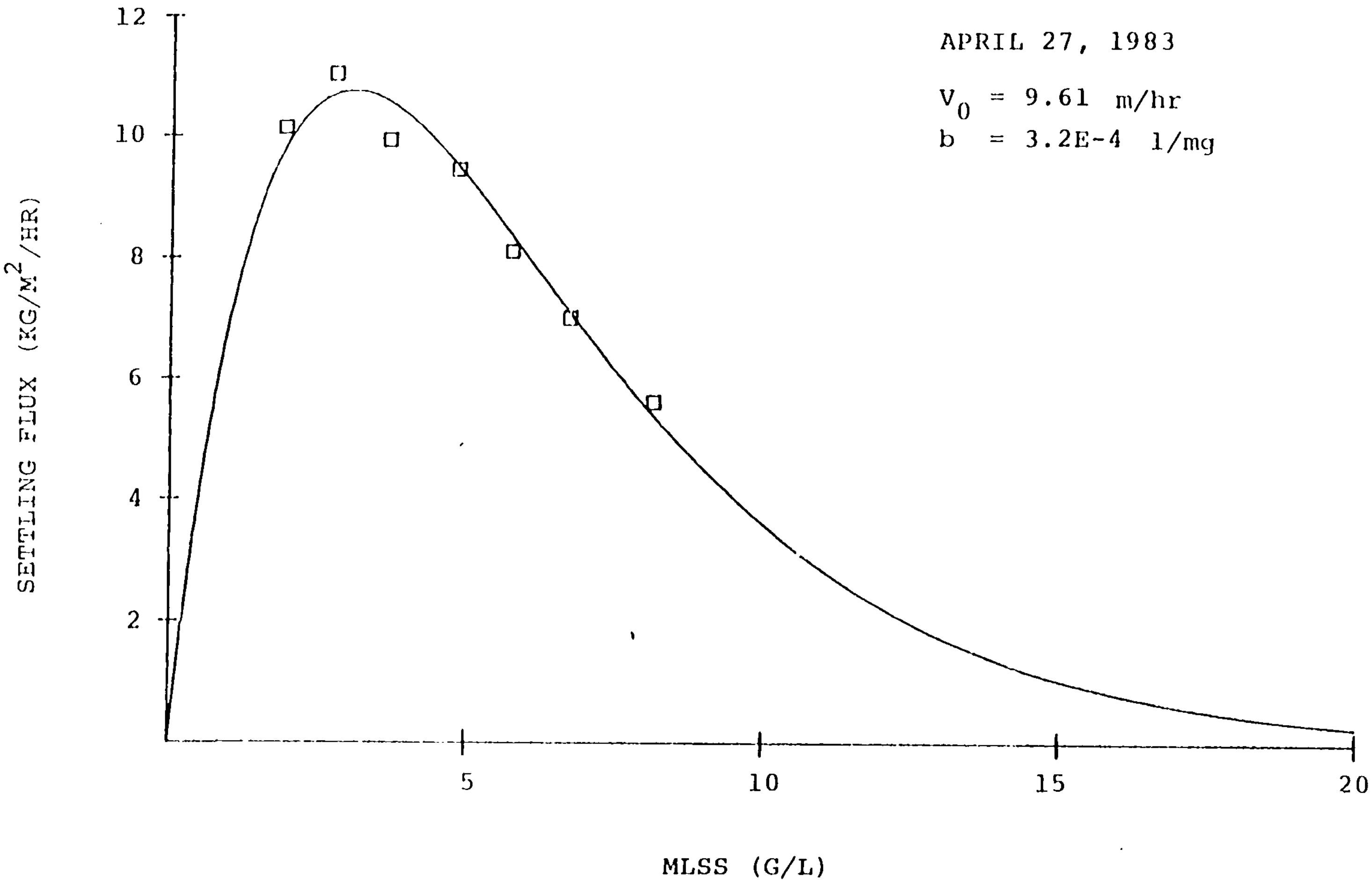

FIGURE 3.12 SETTLING FLUX CURVE

$$Y = X \cdot P \tag{3.2}$$

$$P = [X^T X]^{-1} \cdot X^T \cdot Y \tag{3.3}$$

where:

Y = n x 1 observations,

X = n x p known measurements, observations, etc.,

P = p x 1 model parameters.

X^T = transpose of matrix X,

$[\ ]^{-1}$ = inverse function.

Simplex Optimization

When it was not easy to express the parameter estimation problem in a linear manner, a Simplex optimization technique (103) was used to estimate the model parameters. The error function evaluated by the routine was the square of the difference between observed and simulated data.

The Simplex technique used in this investigation was an unconstrained, sequential search technique using a variable size simplex as developed by Nelder and Mead (83). For an optimization with k factors, k + 1 responses are required to start the procedure. Only one new response is required for each iteration thereafter. The Simplex technique is conceptually simple yet compares favorably (83) with other optimization techniques such as that of Powell (97).

The Simplex optimization technique has been applied in a diversity of scientific and engineering fields. Nelder and Mead (83) used the technique for function minimization. Deming and Morgan (33) surveyed its use in analytical chemistry. Hanashima, at al. (58) used it to estimate flow parameters in their model of a semiaerobic landfill.

It should be noted that this Simplex method is considerably different than the linear programming technique of the same name.

IV. MODEL DEVELOPMENT AND IDENTIFICATION

Several different models were developed during the course of this research. The hydraulic model was developed from well known physical phenomena. Although not formally validated, it did qualitatively describe the flows observed at the plant. Its purpose is to predict the results of hydraulic forcings with time constants on the order of minutes.

The reactor mixing model was selected by comparing the results of tracer tests with several different theoretical models from the literature. Its purpose is to predict suspended solids concentrations in the reactors for use with the settler model and for control purposes.

The settler model is composed of two submodels, these being clarification and thickening models. The clarification model was selected from several models in the literature by comparing experimental results with the results predicted by the different models. A thickening model was selected from the several available on the basis of its ability to describe the phenomena observed in practice. The adequacy of the thickening model was evaluated by comparing predictions of the model with experimental results. Its purpose is to predict the return sludge concentration and the amount of solids stored in the settler.

Reactor Mixing Model

A series of tracer tests were run to delineate a reactor mixing model appropriate for the Sagemont plant. Since the aeration basins are physically segmented into several basins, individual tracer tests were run on each. An impulse of Rhodamine WT dye was injected at the head of each basin, and the concentration at the exit was sampled with an automatic sampler at fixed time intervals. After letting the sample temperature stabilize as described in section III, the dye concentration was determined using with a fluorescence spectrophotometer.

Table 4.1 and Figures 4.1 through 4.4 show the results of these tests (see Figure 3.1 for basin numbering). The figures are presented in terms of normalized concentration and time and show the experimental data as small boxes. Since influent flow rate could not be held constant during the tests, the time variable is approximated by the amount of flow from the time of the injection divided by the reactor volume.

As shown in Table 4.1, the dispersion numbers for the aeration basins are all greater than 0.08 when analyzed as open vessels. Since the dispersion model is best suited to dispersion numbers less than this (74), the dispersion model was not chosen. Plots of the open vessel dispersion curves are shown in Figures 4.1(a) through 4.4(a). While it might be argued that aeration basins are better approximated by closed reactors, the dispersion numbers based upon this assumption are even greater than those for the open vessel. Again, the mixing intensity measured in these basins appears to be greater than the range of applicability of the dispersion model.

TABLE 4.1

TRACER TEST RESULTS

PARAMETER	BASIN 1	BASIN 2	BASIN 3	BASIN 4
Area, C Curve	67.13	56.02	35.37	46.45
Mean Time	0.741	0.9556	1.1155	1.0092
Variance	0.1977	0.6461	0.9938	0.2226
Variance/Mean^2	0.3600	0.6929	0.7444	0.2186
Dispersion Number				
(Closed Vessel)	0.2340	0.8219	1.0428	0.1249
(Open Vessel)	0.1212	0.1947	0.2047	0.08225
Number of CSTR's	2.78	1.44	1.34	4.57

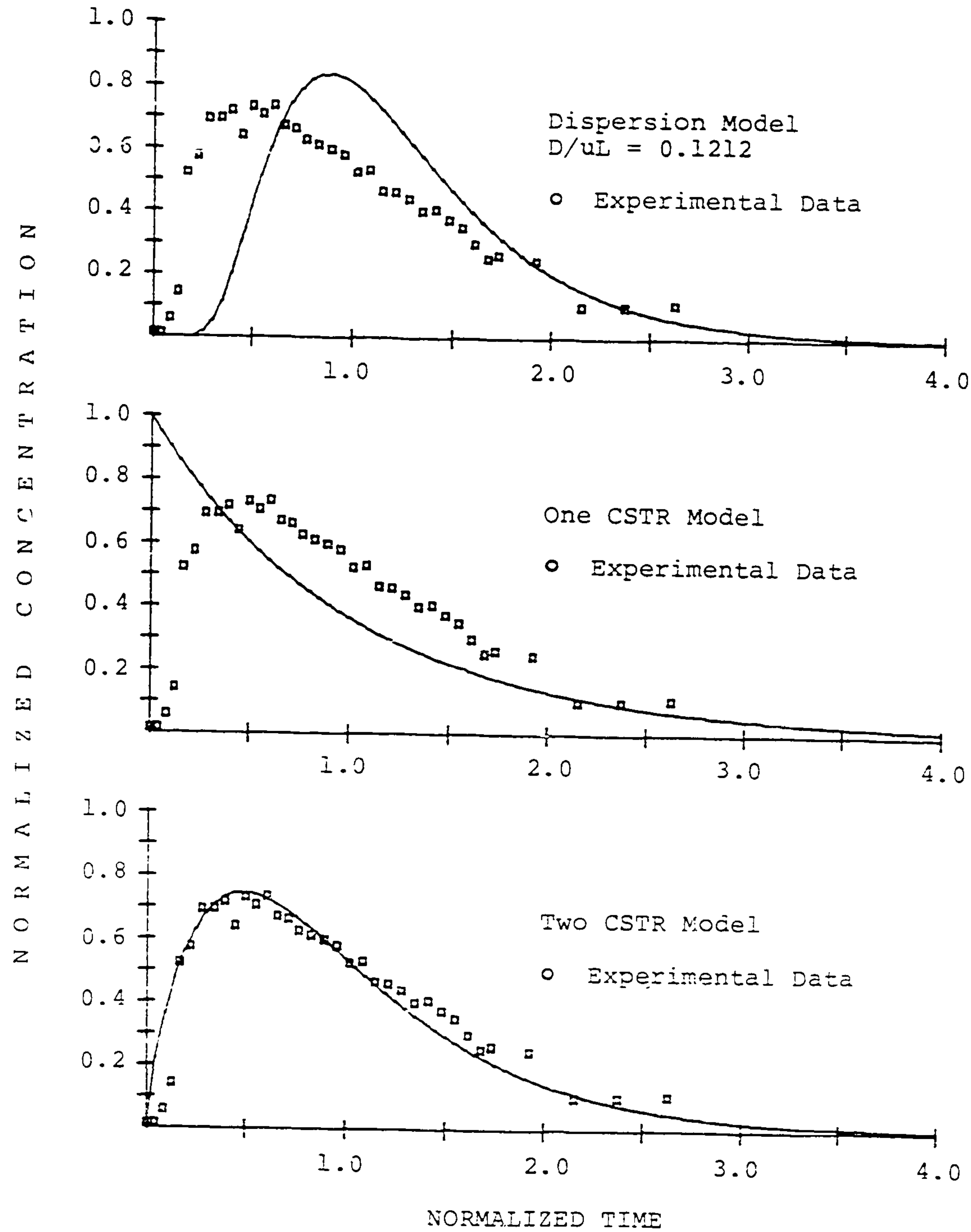

FIGURE 4.1 TRACER TEST - AERATION BASIN 1

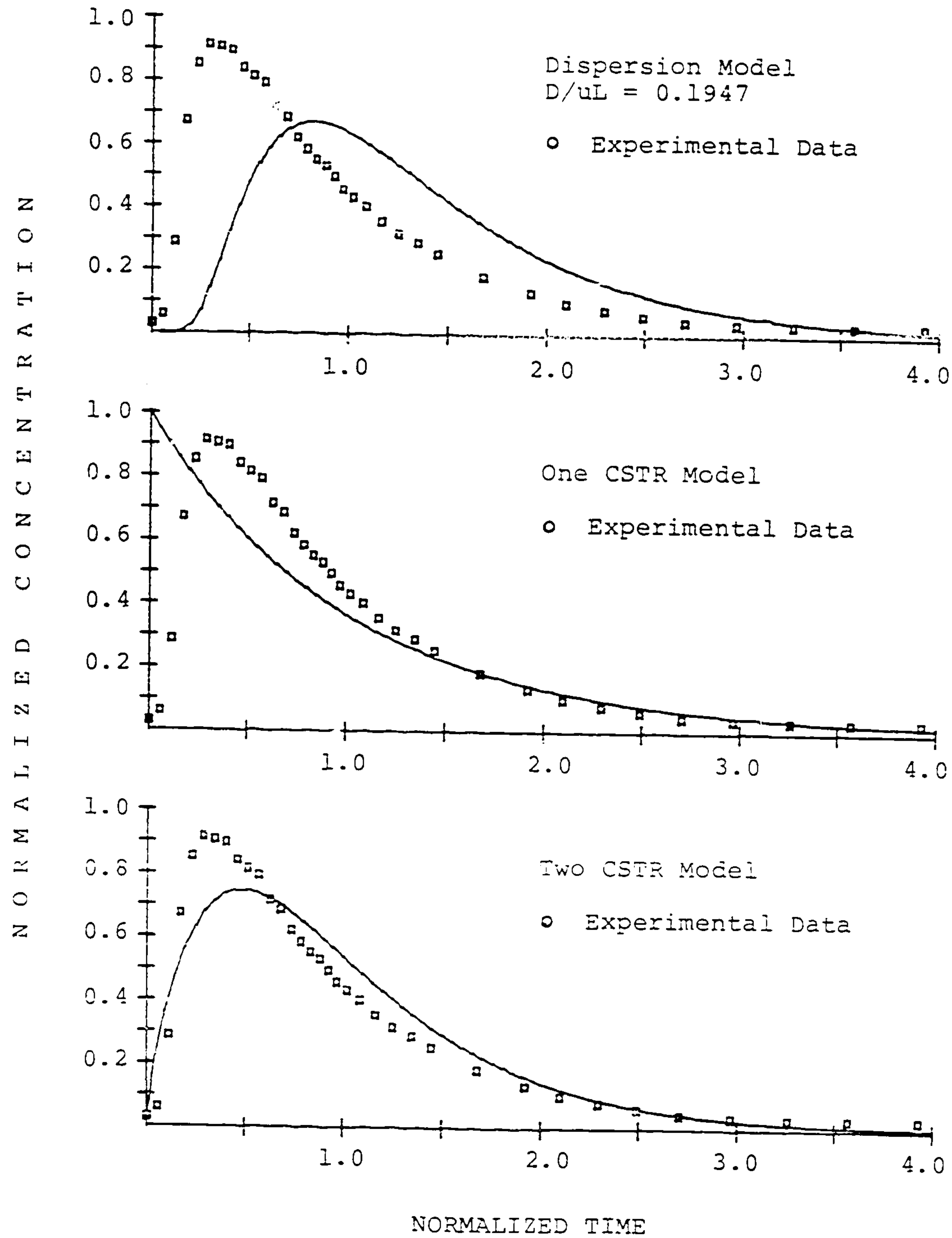

FIGURE 4.2 TRACER TEST - AERATION BASIN 2

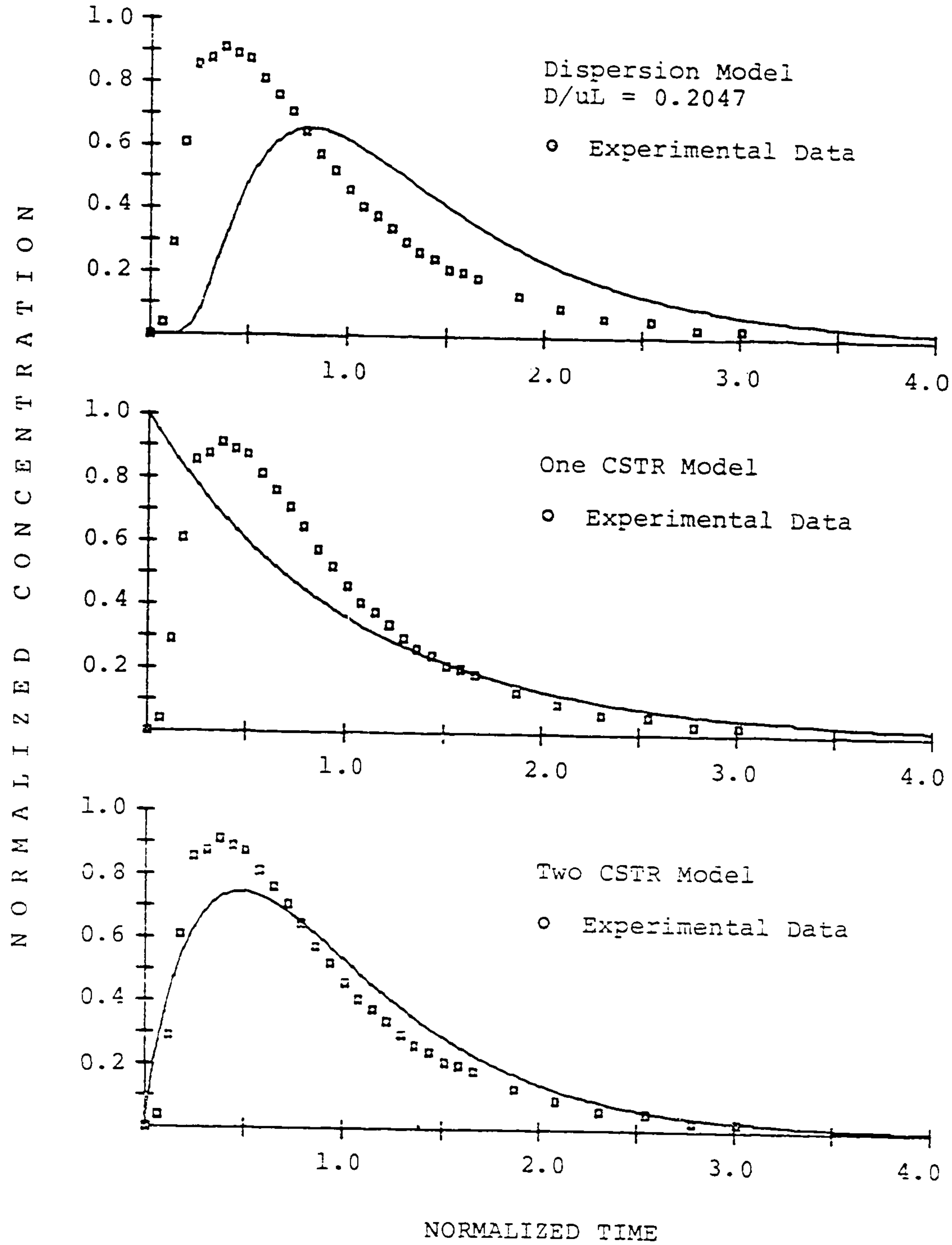

FIGURE 4.3 TRACER TEST - AERATION BASIN 3

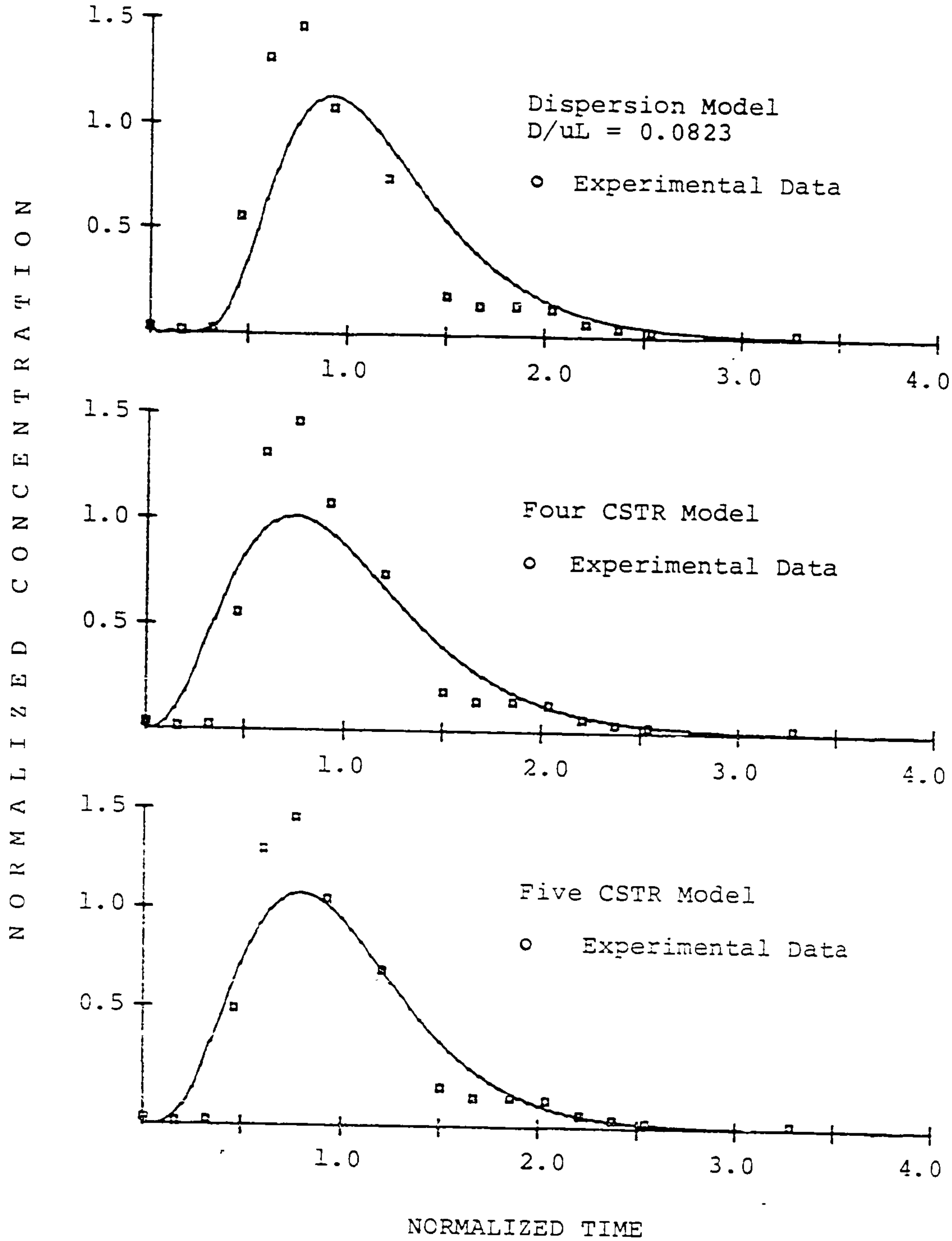

FIGURE 4.4 TRACER TEST - AERATION BASIN 4

The dispersion analysis, however, does give a good estimate of the number of completely mixed reactors in series that it takes to approximate the basins (see Table 4.1). Figures 4.1(b) through 4.4(b) and 4.1(c) through 4.4(c) show the tanks-in-series approximations for the tracer tests. For basin one, two CSTRs is a very good approximation. For basins two and three, the data lies between one and two CSTRs in series. Two CSTRs were chosen so that the model has the same form as the experimental data. For basin four, five CSTRs again provide a reasonable fit.

Tracer tests were not performed on the return activated sludge channels. Therefore, these channels were assumed to have the same mixing characteristics as the parallel aeration basins of the same length. Thus, return sludge channel 1, which parallels aeration basin 4, was assumed to be characterized by five CSTRs in series. Return sludge channel 2, which parallels aeration basin 3, was assumed to be equivalent to two CSTRs in series. Table 4.2 gives a summary of the reactor mixing model adapted for the Sagemont WWTF.

Multiparameter models were not evaluated since the single parameter tanks-in-series model adequately fit the data. Although it is realized that multiparameter models could better describe the data, the tanks-in-series model is adequate for the purpose of this study. Simulation results presented in a later section compare favorably with experimental data. The tanks-in-series model also has the computational advantage of involving only ordinary differential equations which can be easily solved on a small process control computer.

TABLE 4.2

REACTOR MIXING MODEL SUMMARY

Basin	Number of CSTRs	Volume of Each (Cu M)	Total Volume (Cu M)
Basin 1	2	785	1,570
Channel	1	1,032	1,032
Basin 2	2	688	1,376
Basin 3	2	382	764
Basin 4[1]	5	101	505
RAS Channel 1	5	101	505
RAS Channel 2	2	191	382

[1] Volumes for basin 4 are based upon 50% useful volume due to grit deposits.

<u>Hydraulic Reactor Model</u>

The hydraulic reactor model dynamically predicts all flows between reactors and the height of water in each reactor. This is accomplished by deriving a mass balance equation for each reactor in the form of Equation [4.1].

$$\frac{d(H)}{dt} = \frac{F_{IN} - F_{OUT}}{AREA} \qquad [4.1]$$

where:

H = height of water in reactor (L),

F_{IN} = total flow into reactor (L^3/T),

F_{OUT} = total flow out of reactor (L^3/T),

AREA = surface area of reactor (L^2).

The flows between reactors depend upon the outlet type and the relative heights of water in the adjacent reactors. Three types of outlets and an open end between reactors have been modeled for the Sagemont plant. Details of these are shown in Table 4.3 and Figure 4.5.

The equation for flow over a rectangular weir with end contractions is shown as Equation [4.2].

$$F = 3.33\ (L-0.2H)\ H^{3/2} \qquad [4.2]$$

where:

F = flow (cfs),

L = length of weir (ft),

H = height of water (ft).

TABLE 4.3

HYDRAULIC REACTOR MODEL SUMMARY

Basin	Area (Sq Ft)	Outlet Type	Outlet Constants
Bar Screen	420	rect. Weir	L=20 ft
Basin 1	3,735	slot	K=0.5 A=varies
Channel	2,470	2 sluice gates	K=0.5 A=18 Sq Ft
Basin 2	3,330	2 sluice gates	K=0.5 A=24.5 Sq Ft
Basin 3	1,847	open end	
Basin 4	4,500	sluice gate & pipe	K=1.5 A=9.62 Sq Ft
Clarifier	7,854	V-notch weirs	n=665
RAS Channel 1	2,180	sluice gate & pipe	K=1.0 A=12.57 Sq Ft
RAS Channel 2	923	sluice gate	K=0.5 A=16.0 Sq Ft
Chlorine Inlet	280	2 sluice gates	K=0.5 A=18.0 Sq Ft
Chlorine Basin	3,120	V-notch weirs	n=350
Chlorine Outlet	666	V-notch weir	n=1

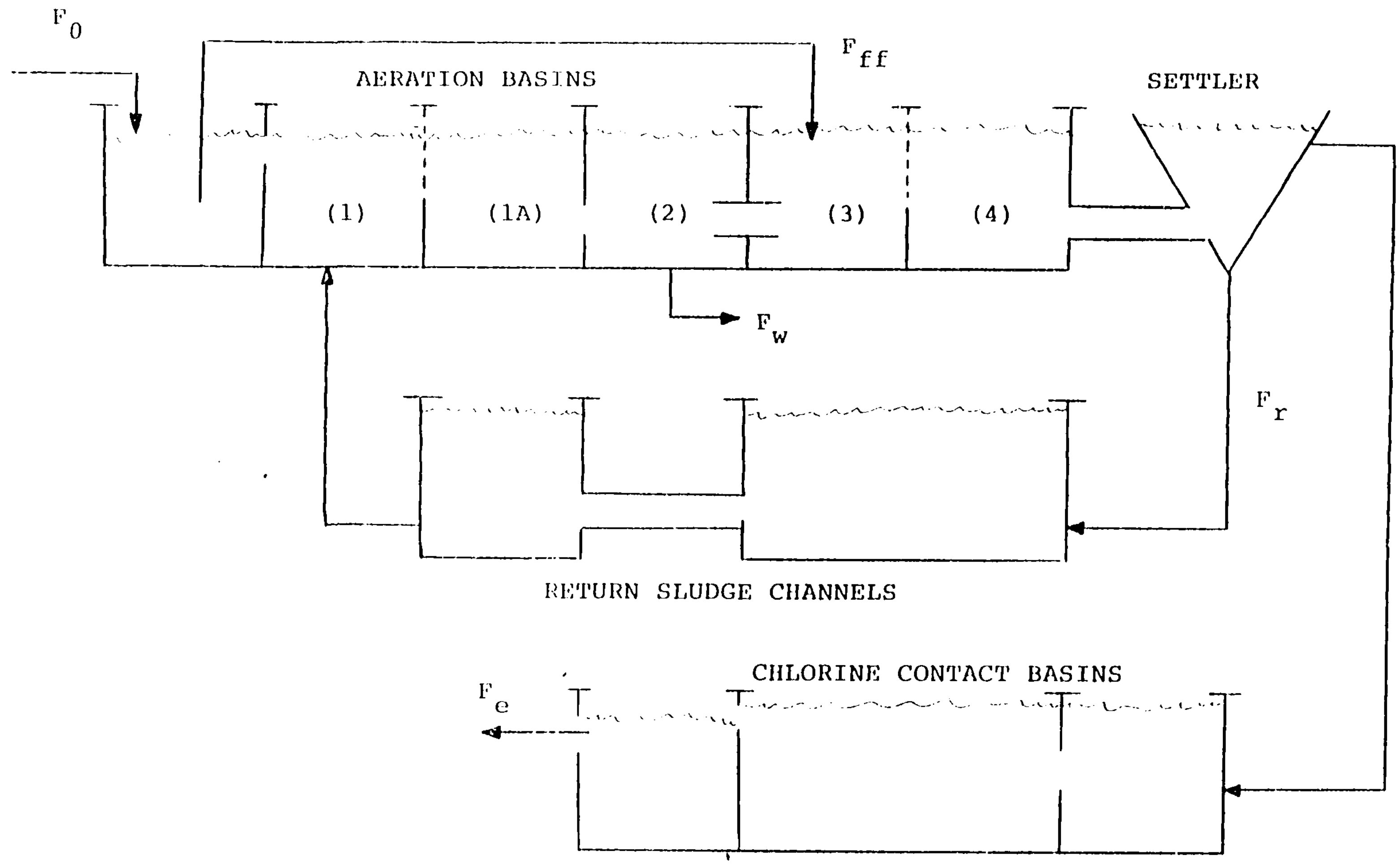

FIGURE 4.5 HYDRAULIC REACTOR MODEL SCHEMATIC

The equation for flow over a 90 degree (V-notch) weir is shown as Equation [4.3].

$$F = 2.5\ H^{5/2} \qquad [4.3]$$

Head loss through an opening or sluice gate is proportional to the velocity squared as shown in Equation [4.4].

$$H = K \cdot \frac{V^2}{2g} \qquad [4.4]$$

where:

V = velocity of flow (L/T),

g = gravitational constant (L/T^2),

K = head loss coefficient.

Equation [4.4] can be rearranged to solve for flow ($F=V \cdot A$) as shown in Equation [4.5].

$$F = A \cdot (2g/K)^{1/2} \cdot (H)^{1/2} \qquad [4.5]$$

where:

A = area of opening (L^2).

Flow through a pressurized pipe is often expressed with the Hazen-Williams equation, [4.6].

$$F = 1.318 \cdot A \cdot C_H \cdot R^{2/3} \cdot S^{1/2} \qquad [4.6]$$

where:

C_H = coefficient,

R = hydraulic radius (L),

S = slope of the energy line.

Equation [4.6] can also be expressed in a form similar to Equation [4.5] with an equivalent coefficient, K. In this way, head loss through an opening and pipe combination can be calculated with one coefficient.

There is no gate between basins 3 and 4, just a large opening. Therefore, there is no equation for flow out of basin 3. In the model, the level in each basin is assumed to be the same. In a dynamic system, this also means that the derivatives of the heights in each basin (Equations [4.7] and [4.8]) must also be equal. Equations [4.7] and [4.8] can then be combined to solve for the flow out of basin 3 as shown in Equation [4.9].

$$\frac{d(H_3)}{dt} = \frac{(F_2 + F_{ff} - F_3)}{A_3} \qquad [4.7]$$

$$\frac{d(H_4)}{dt} = \frac{(F_3 - F_4)}{A_4} \qquad [4.8]$$

$$F_3 = \frac{A_3 \cdot F_4 + A_4 \cdot (F_2 + F_{ff})}{A_3 + A_4} \qquad [4.9]$$

Due to the low velocities through the aeration basins, head loss through any aeration basin is insignificant.

Full-Scale Experimental Data Summary

Summaries of the eight full-scale experiments performed during this investigation are presented herein. The data from these experiments were used to identify the clarification and thickening models and estimate model parameters. Each plot shows eight hours of one minute average values for each of five on-line measurements. These measurements included effluent flow rate (Fe), recycle flow rate (Fr), effluent suspended solids concentration (Xe), mixed liquor suspended solids concentration in the last aeration basin (X4), and return sludge solids concentration (Xr). Since the controlled portion of each test typically lasted three to five hours, each plot also shows one or more hours of data before and after the experiment. Note that the scales for effluent flow rate, MLSS, and Xe vary between figures.

Experiments

Experiment 1. In experiment 1 (Figure 4.6) the recycle flow rate was manipulated in steps of up to 2.5 MGD (300 gpd/square foot) to investigate the influence of this parameter on effluent suspended solids concentration. Influent flow rate was not manipulated.

Experiment 2. The purpose of experiment 2 (Figure 4.7) was to study possible "overshoots" of effluent suspended solids caused by step increases in flow as observed by Chapman (20) in his pilot study. In this experiment the influent flow was held at three (near) constant rates with step changes between each level. Additionally, the recycle flow rate was decreased to a low value to overload the settler. This resulted in a decreasing MLSS concentration feed to the settler.

Experiment 3. The purpose of experiment 3 (Figure 4.8) was to determine the affect of sludge blanket level on effluent suspended solids. In this experiment two flow pulses were applied to the settler. Between

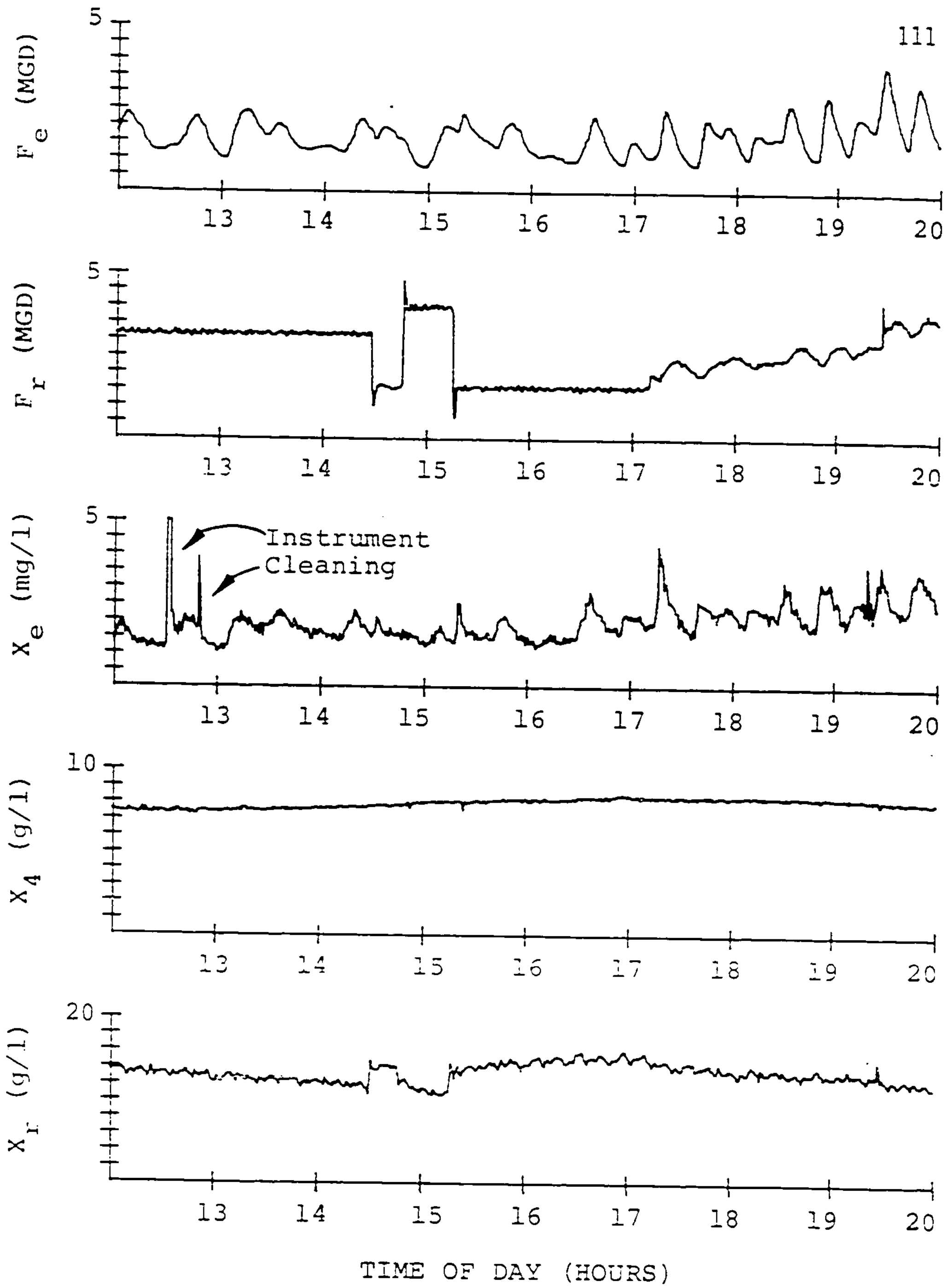

FIGURE 4.6 EXPERIMENT 1 - APRIL 21, 1983

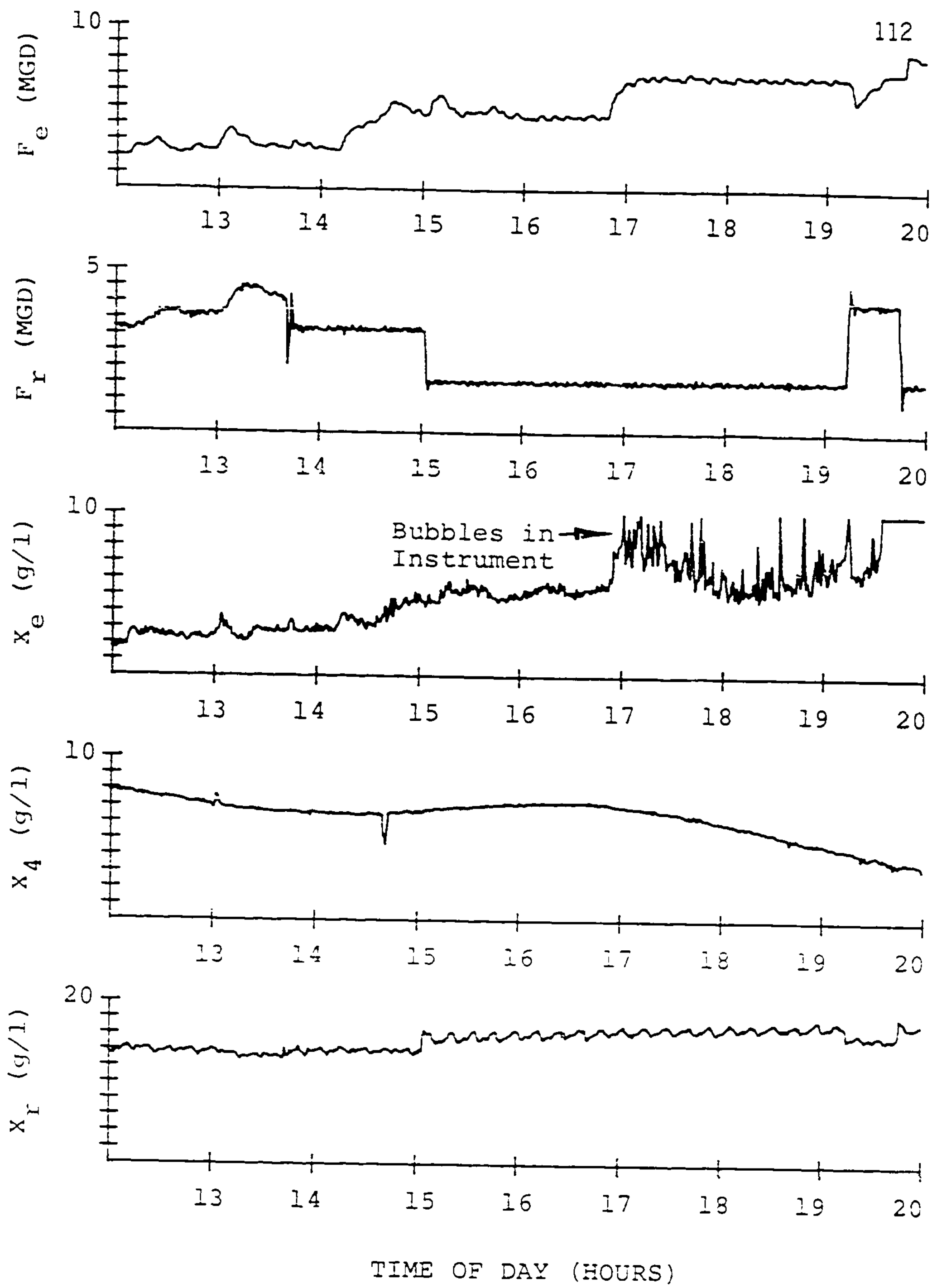

FIGURE 4.7 EXPERIMENT 2 - APRIL 23, 1983

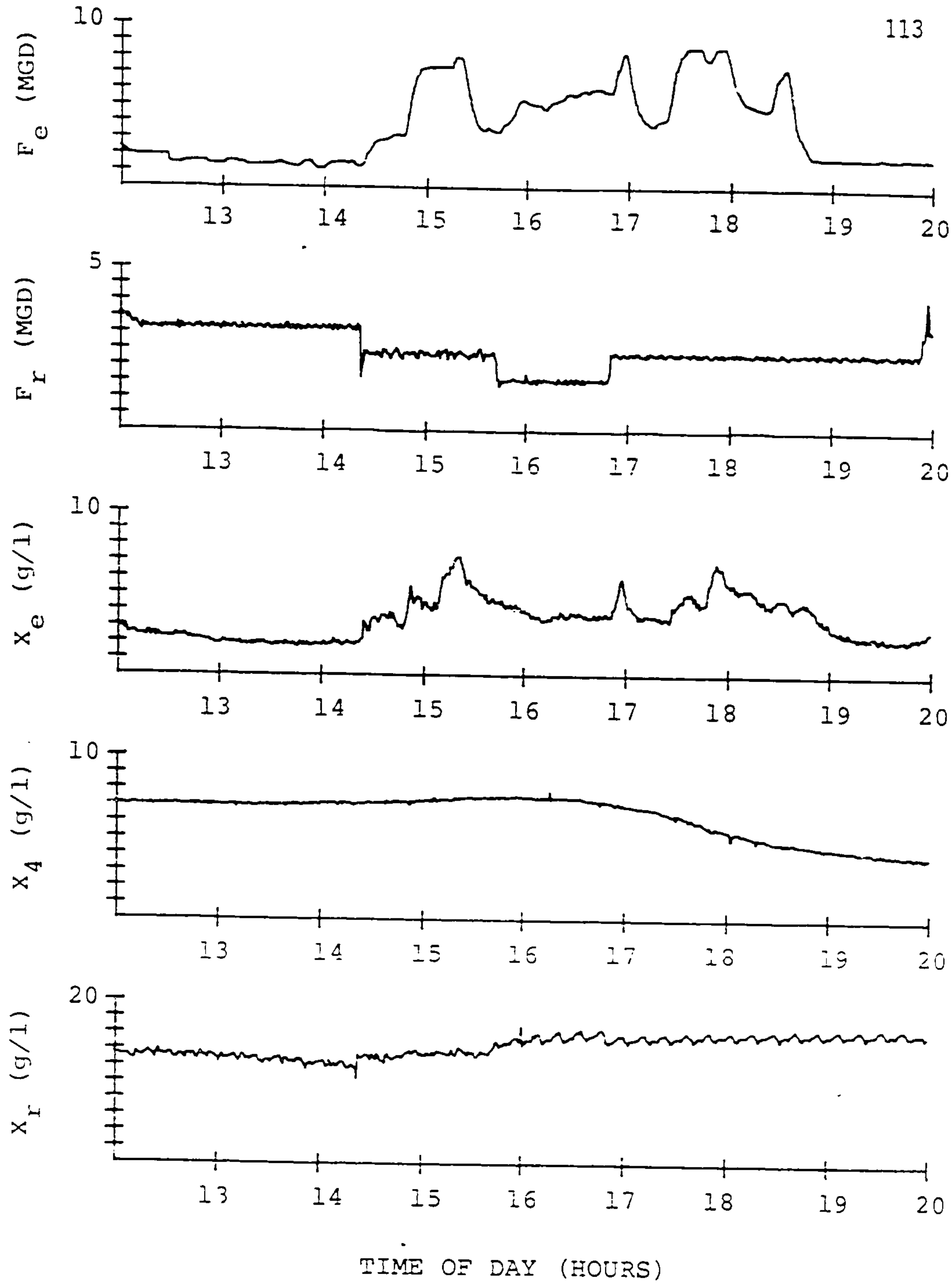

FIGURE 4.8 EXPERIMENT 3 - APRIL 26, 1983

the pulses, the sludge blanket was allowed to build up. Thus, one pulse was applied with a low sludge blanket and the other with a high blanket level.

Experiment 4. Experiment 4 (Figure 4.9) was one of two tests involving the use of the step feed capability of the Sagemont plant. While the influent flow rate was held constant, approximately 85 percent of the influent flow was diverted to the third aeration basin. Near zero level dissolved oxygen levels were measured in aeration basins 3 and 4. The original purpose of the test was to observe effluent suspended solids at a constant flow rate and changing MLSS concentration.

Experiment 5. Experiment 5 (Figure 4.10) was similar to experiment 4 with the exceptions that the dissolved oxygen concentration in the aeration basins was held at a level of 2.5 mg/l or greater and the portion of feed forward flow was approximately 65 percent of the influent flow rate.

Experiment 6. The purpose of experiment 6 (Figure 4.11) was to investigate the effect of rate of increase of influent flow rate on effluent suspended solids. Two flow pulses with rates of increase of 0.4 MGD/minute and 0.2 MGD/minute were applied to the settler.

Experiment 7. Experiment 7 (Figure 4.12) was similar to experiment 6. The rate of increase of flow for this test was chosen to be 0.1 MGD/minute. This long flow pulse overloaded the settler allowing the observation of an increasing sludge blanket during the pulse and a decreasing blanket level after the pulse.

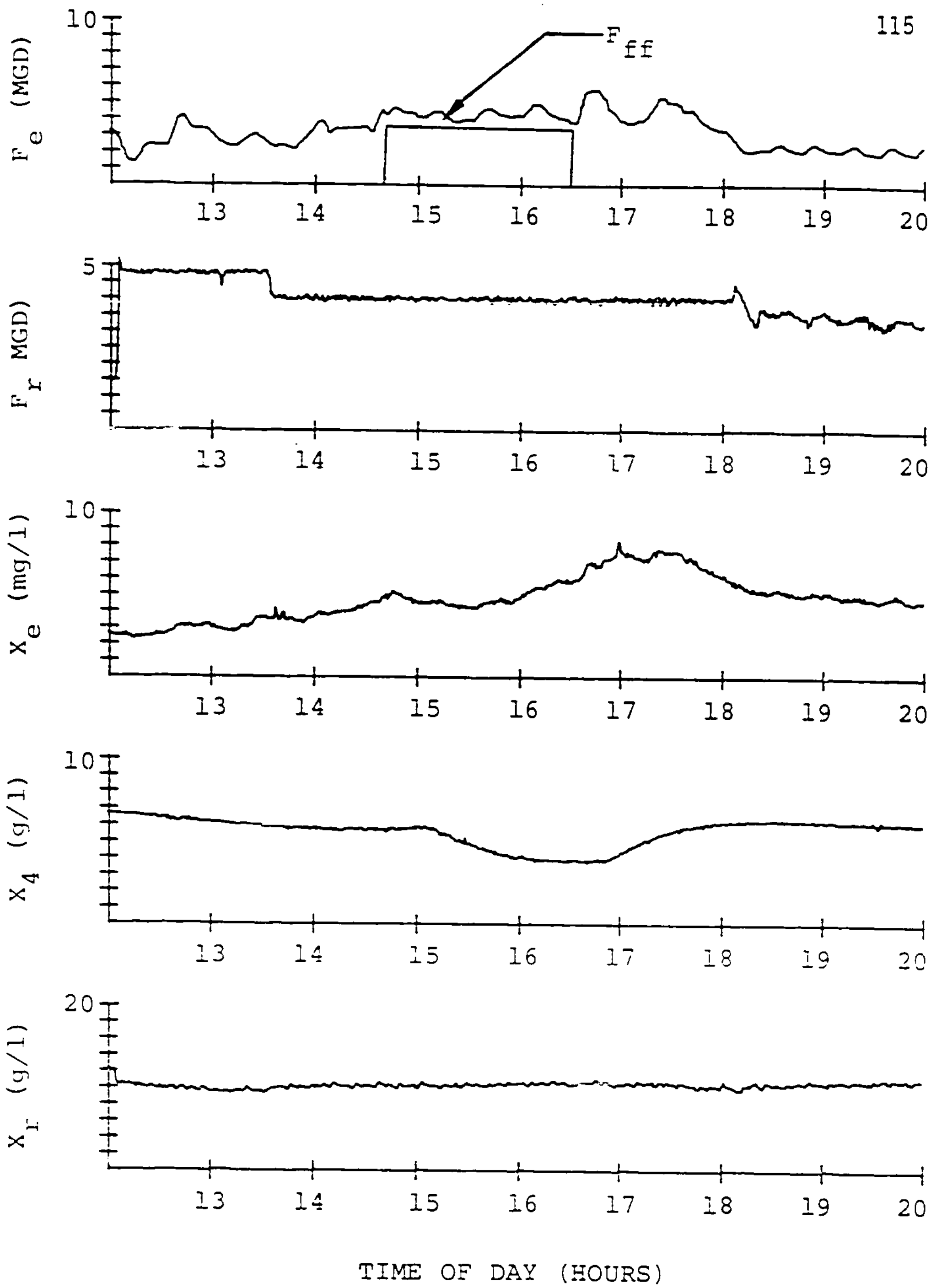

FIGURE 4.9 EXPERIMENT 4 - JUNE 25, 1983

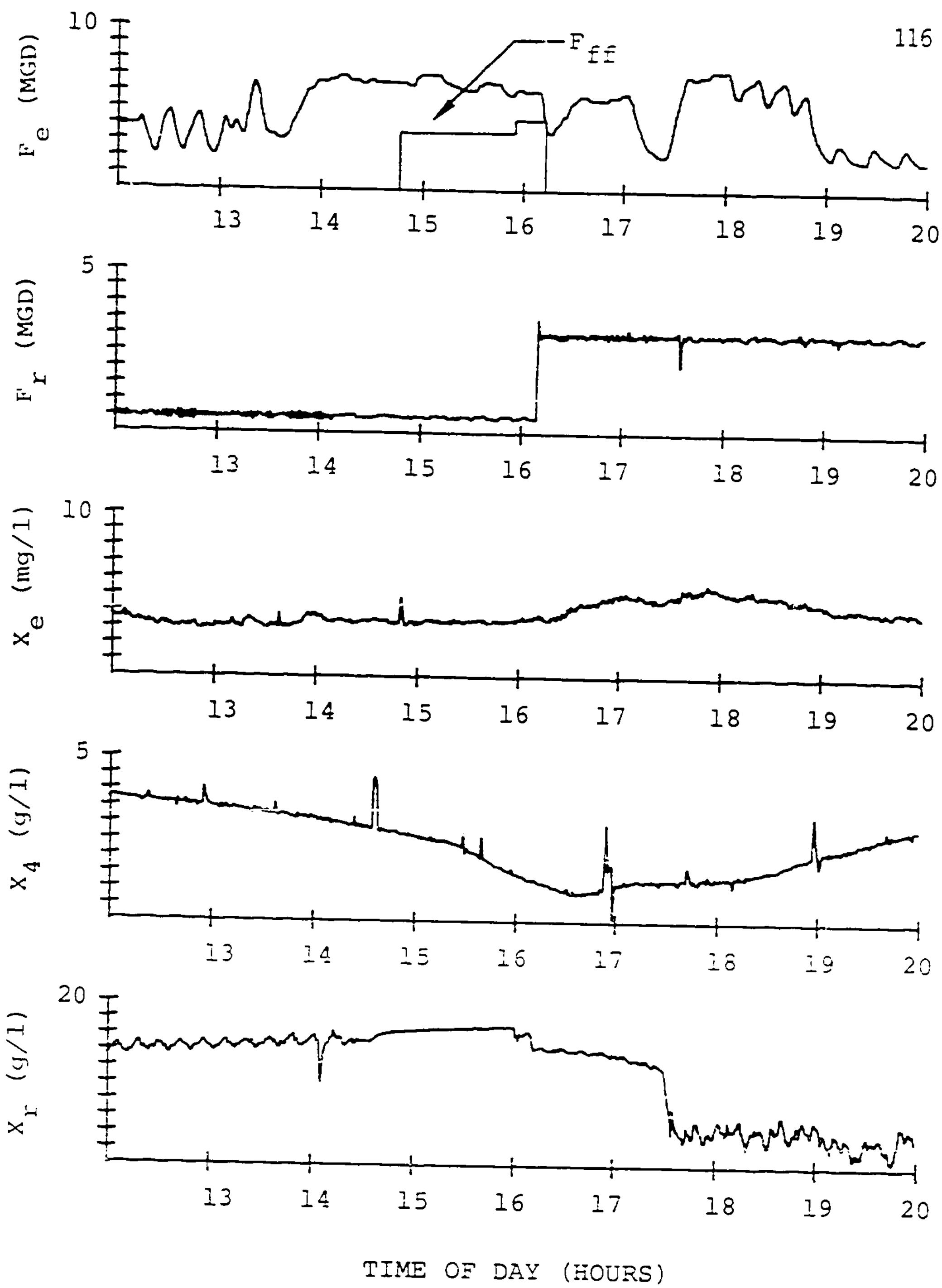

FIGURE 4.10 EXPERIMENT 5 - JULY 27, 1983

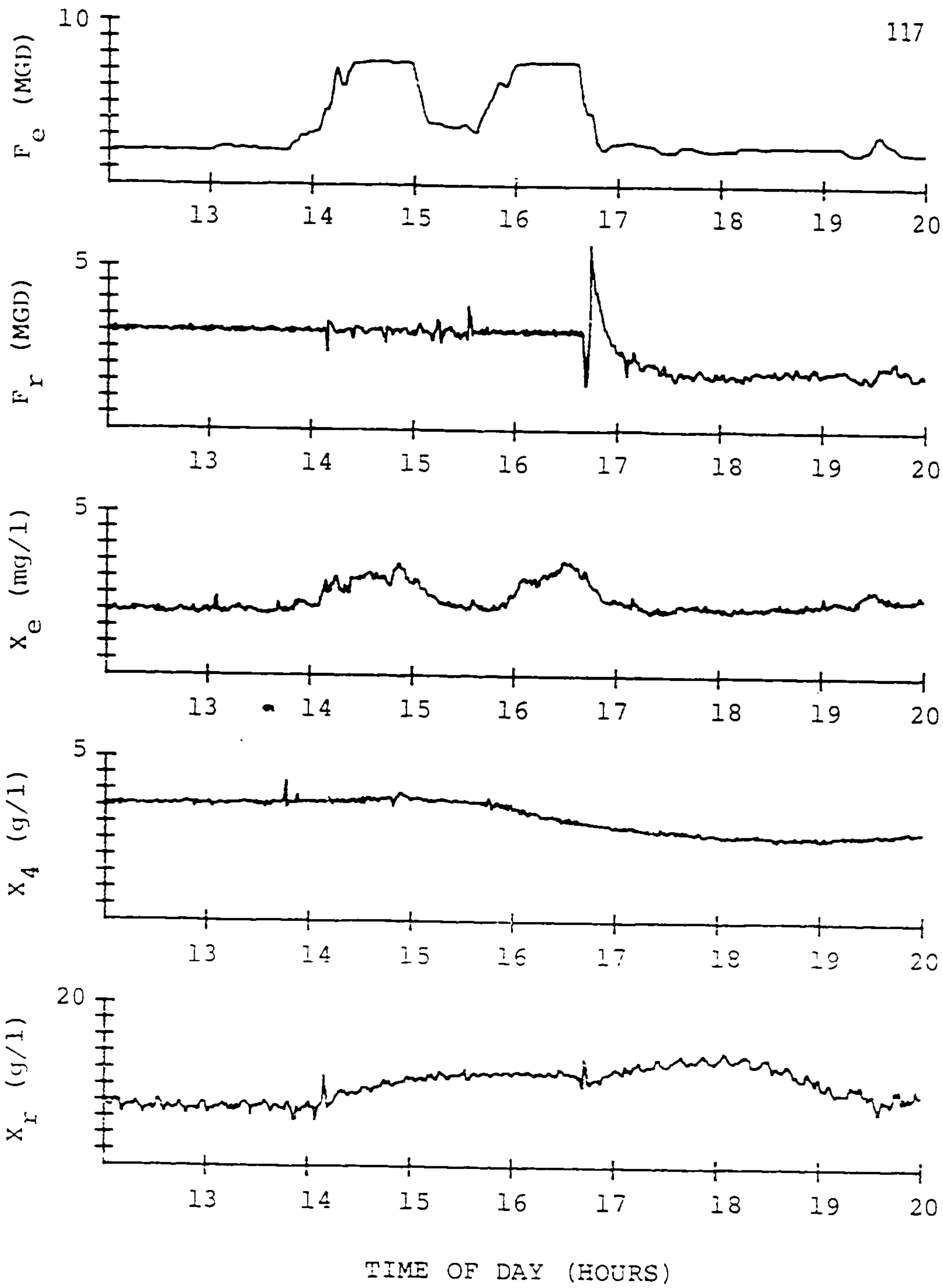

FIGURE 4.11 EXPERIMENT 6 - AUGUST 2, 1983

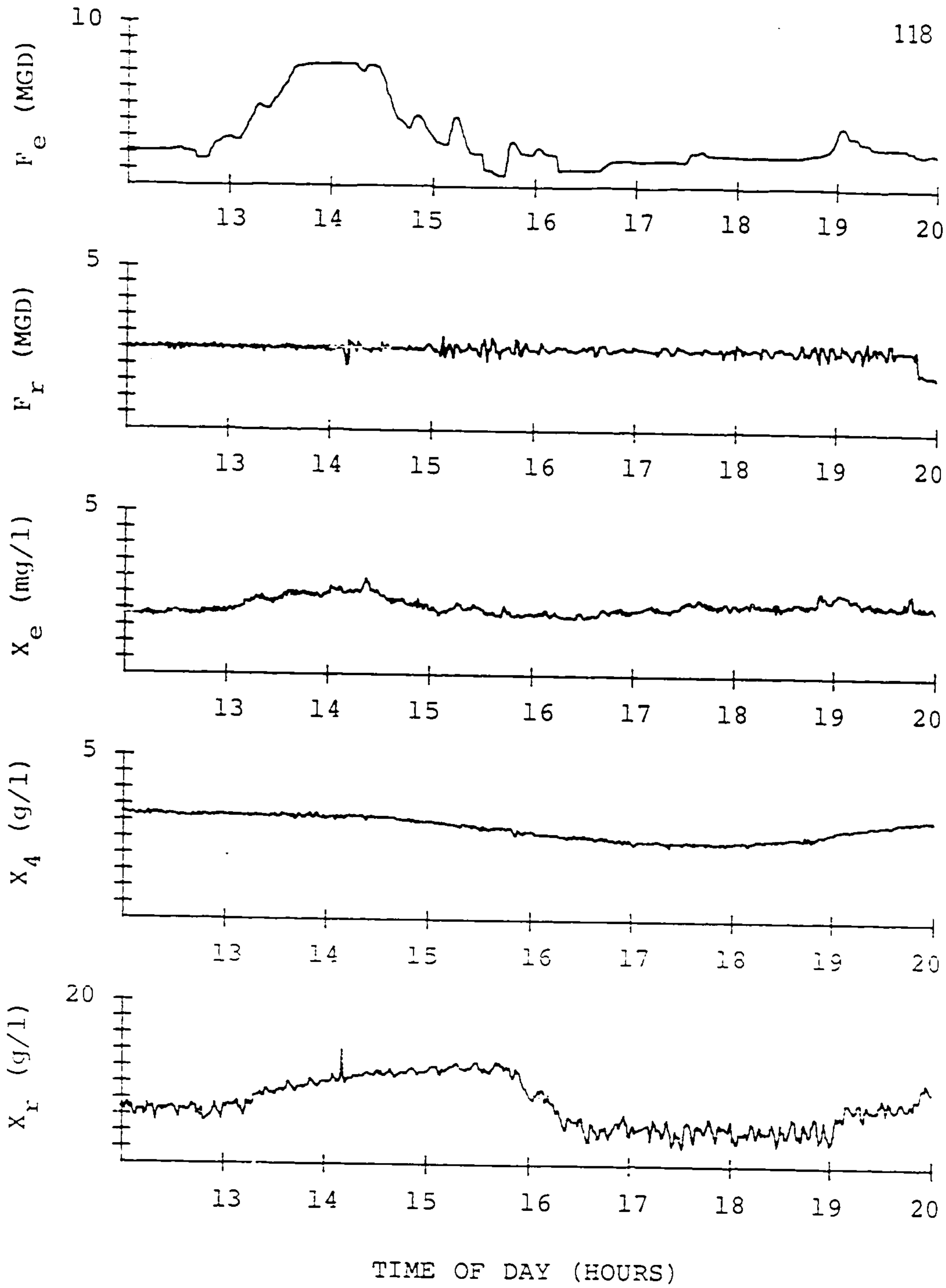

FIGURE 4.12 EXPERIMENT 7 - AUGUST 3, 1983

Experiment 8. Experiment 8 (Figure 4.13) was essentially repetition of experiment 6. Data after the second pulse were not available due to operator diversion of the effluent flow to an empty basin.

Observed Parameters

Effluent Flow Rate. In experiment 1 (Figure 4.6), the flow pattern was not artificially modified. It shows the natural variations induced by multiple lift stations pumping to the plant in combination with the hydraulic dampening capacity of the basins. It should be noted that two off-site lift stations pump directly to the Sagemont plant while several others go to an on-site lift station. It was not possible to control the off-site lift stations. Fortunately, the contributions from these two sources usually amounted to less than twenty percent of the total inflow. During this test, the variable frequency pump operating with a constant level algorithm was on-line.

In the remaining experiments, the influent flow was controlled with a local PI control loop on the programmable controller (PC) in the lift station. Treated effluent was drained from the off-line clarifier and chlorine contact basin back to the lift station to provide the required water. Drain valves on the basins were manually adjusted to maintain an adequate supply. The wet well level was constantly monitored to avoid overflowing or emptying.

For the tests from April through July (Figures 4.6 to 4.10), the flow set point was adjusted manually on the PC loop access module. For the August tests (Figures 4.11 to 4.13), software was developed to

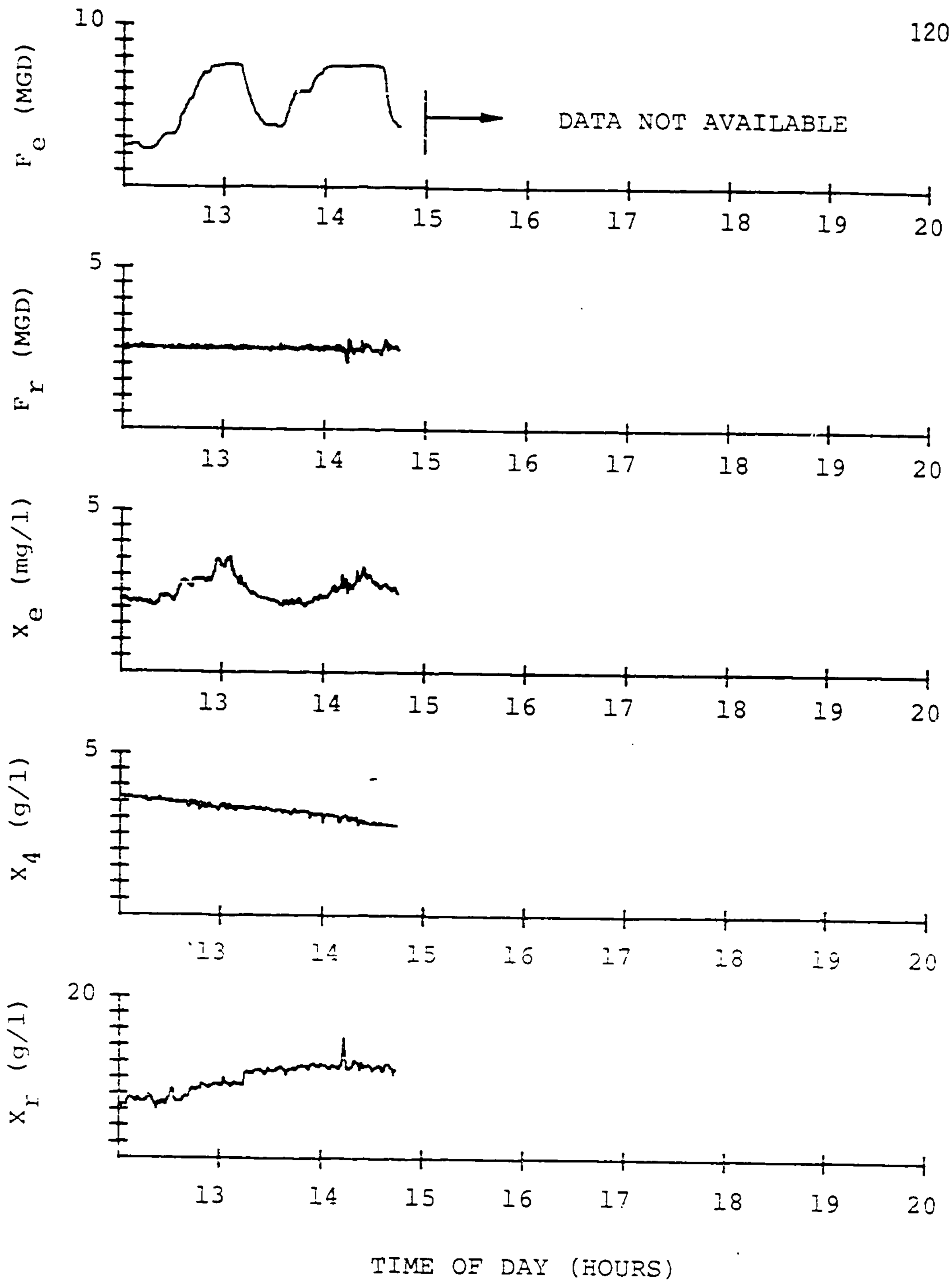

FIGURE 4.13 EXPERIMENT 8 - AUGUST 7, 1983

automatically download a series of flow set points from a table on the main computer. The set point could be changed at a rate of once per minute. This allowed arbitrary rates of change of the flow rate.

Return Sludge Flow Rate. The return sludge (or recycle) flow rate was manually set to a constant value or proportional to an exponentially filtered value of the effluent flow. Note that many of the plots show a change of control strategy (constant or proportional) before or after the controlled portion of the test.

The recycle flow rate was controlled by a single PI control loop on the PC. A simple algorithm in the PC was used to decide whether one or two of the recycle airlifts pumps were needed. When the controller output exceeded 80 percent, two airlifts were used. Both remained on-line until the controller output fell below 35 percent. This logic lead to the overshoots and undershoots of recycle flow shown on several of the figures. These deviations typically lasted less than two minutes and were not considered detrimental to the experiments. Additional PC logic alternated the two recycle airlifts every four hours to prevent clogging due to infrequent use.

Effluent Suspended Solids Concentration. The effluent solids observed at the Sagemont plant varied between 1.5 and 8.0 mg/l during the tests reported herein. In fact, the highest concentration observed under normal operating conditions was about 15 mg/l. These values are generally considered low for the activated sludge process. Although there is no single reason for these low values, several factors probably contribute. These factors include a long sludge age (10-40 days), a long hydraulic detention time (12 hours), high water temperatures (75-90 degrees F), low settler overflow rates, and deep settlers.

In experiment 1 (Figure 4.6), several peaks due to instrument cleaning are shown. These points were eliminated from the data before estimating parameters.

In experiment 2 (Figure 4.7), a deficiency of the effluent solids sampling system was observed. When the flow rate increased above 7 MGD (26.5 M l/day), air bubbles were entrained in the sample stream. These bubbles resulted in high, erratic readings from the instrument. The data from this portion of the plot were not used in the analysis. A debubbler was installed in the sample stream for the tests following this incident.

Mixed Liquor Suspended Solids Concentration. The MLSS concentration observed at the Sagemont plant varied from the 7,000 - 8,000 mg/l range for the tests from April to June (Figures 4.6 to 4.9) and from the 3,000 - 4,000 mg/l range for the July and August tests (Figures 4.10 to 4.13). This variation was accomplished by changing the sludge age from 40 days to 15 days. Concentrations as low as 900 mg/l were recorded during one of the feed forward experiments (Figure 4.10).

The MLSS instrument ordinarily gave a steady reading with only 1 to 2 percent noise. However at times, rags and other pieces of trash obstructed the optical surfaces of the solids probe resulting in erroneous readings. Most of these incidents lasted only a few minutes as shown in Figures 4.7 and 4.10. It is interesting to note that these deviations can be both positive and negative. Apparently rags and other opaque objects increase the reading while transparent objects (plastic bags) decrease the reading.

Return Activated Sludge Concentration. The return activated sludge concentrations varied from as high as 16,000 mg/l to as low as 4,000 mg/l. Additionally each plot shows a periodic variation with a time period of minutes. This phenomenon will be discussed further in a later section.

Clarification Model

A series of experiments were conducted to delineate which (if any) of the models from the literature adequately described the clarification function of the settler. In order to resolve some of the conflicting observations reported by other researchers, the experiments were designed to answer several key questions.

1) What influence does the recycle flow rate have on effluent solids? Is the effluent flow rate or total flow rate of more importance?
2) What are the effects of MLSS concentration? Do higher MLSS concentrations increase or decrease effluent solids?
3) What are the effects of a high sludge blanket? Does it enhance or decrease solids capture or have little effect?
4) What are the effects of hydraulic transients? How long does it take for the effects of hydraulic transients to die out?

Effects of Recycle Flow Rate

Several experiments were conducted in which step changes in the return sludge flow rate were introduced. Results from these three experiments are summarized in Figures 4.6 (April 21, 1983), Figure 4.7 (April 23, 1983), and Figure 4.8 (April 26, 1983). Of these experiments, the one from April 21 most clearly shows the effects of recycle flow rate. It should be noted that this experiment had the largest changes in recycle flow rate both in absolute value and as a portion of the inflow (recycle ratio).

Two models were evaluated to quantify the effects of recycle flow rate on effluent solids. Both are similar to the basic Pflanz (95) model. They are:

$$Xe = K_1 + K_2 \cdot Fe \tag{4.10}$$

$$Xe = K_1 + K_2 \cdot (Fe+Fr) \tag{4.11}$$

where:

K_1, K_2 = model parameters,

Fe = effluent flow rate (L^3/T),

Fr = recycle flow rate (L^3/T).

Both of the above equations differ from the Pflanz model by including a flow independent parameter, K_1. This constant varied from 0.5 to 2.0 mg/l. The inclusion of this term is consistant with the observation in this research that Xe always remain finite (0.5 to 2.0 mg/l) even when low flows (<0.5 MGD) are encountered for several hours. Pflanz's data also showed this type of offset but could be ignored since it was a small fraction of the solids range that he observed (20 to 80 mg/l).

Tables 4.4 and 4.5 show the least square estimates of the parameters for Equations [4.10] and [4.11]. Table 4.6 summarizes these results (and others) by comparing the sum of the squares of the residuals for these data. Clearly the model excluding recycle flow rate (Equation [4.10]) gives a much better fit than the model that include it. This is shown graphically in Figures 4.14 through 4.17 which show the data and predictions of Equations [4.10] and [4.11] for the first two experiments. It is especially apparent in Figures 4.14 and 4.15 (April 21, 1983) that the addition of recycle flow rate as a model input does not add any significant information to the prediction of Xe.

Examination of the parameters in Table 4.4 shows a wide variation of the parameter K_2 between the different experiments. This is especially apparent in comparing the results of the April and August tests. These two sets of experiments were performed at different sludge ages (40 and 15 days, respectively) resulting in substantially different MLSS concentrations. The effect of MLSS concentration is discussed later in this section. The rate at which these parameters vary and techniques for following these changes are discussed in section VII.

In summary, the effluent solids concentration appears to be more of a function of overflow rate than total flow (Fe+Fr) for the clarifier studied. This is in agreement with Pflanz (95). A similar conclusion was also reached by Sorensen (106) while working with pilot plants with recycle ratios ranging from 50 to 200 percent. There is qualitative evidence from other plants (52), however, which suggest that underflow rates greater than 1000 gpd/sq ft (1.7 m/hr) may lead to a degradation

TABLE 4.4

PARAMETERS FOR EQUATION [4.10]

$$Xe = K_1 + K_2 \cdot Fe$$

Date	K_1	K_2	NUMBER DATA POINTS	SUM RESIDUAL SQUARED	AVG ERROR
April 21, 1983	0.78	0.7003	480	71.4	0.39
April 23, 1983	0.16	0.9752	285	95.7	0.58
April 26, 1983	1.68	0.4661	480	230.3	0.69
Aug. 2, 1983	1.74	0.1641	480	22.7	0.22
Aug. 3, 1983	1.79	0.1095	480	22.7	0.22
Aug. 7, 1983	1.81	0.1325	166	12.0	0.27

TABLE 4.5

PARAMETERS FOR EQUATION [4.11]

$$Xe = K_1 + K_2 \cdot (Fe+Fr)$$

Date	K_1	K_2	NUMBER DATA POINTS	SUM RESIDUAL SQUARED	AVG ERROR
April 21, 1983	1.10	0.2010	480	116.9	0.49
April 23, 1983	2.13	0.2513	285	424.0	1.22
April 26, 1983	0.25	0.5072	480	269.1	0.75
Aug. 2, 1983	1.52	0.1303	480	29.6	0.25
Aug. 3, 1983	1.49	0.1147	480	18.8	0.20
Aug. 7, 1983	1.47	0.1330	166	12.0	0.27

TABLE 4.6

SUM OF RESIDUALS SQUARED FOR CLARIFICATION EQUATIONS

Date	NUMBER DATA POINTS	MODEL NUMBER [4.10]	[4.11]	[4.12]	[4.13]	Decay Model[1]
April 21, 1983	480	71.4	116.9	59.9	102.7	66.5
April 23, 1983	285	95.7	424.0	74.0	424.7	36.8
April 26, 1983	480	230.3	269.1	294.5	405.4	131.0
Aug. 2, 1983	480	22.7	29.6	29.1	39.9	19.0
Aug. 3, 1983	480	22.7	18.8	24.8	23.4	21.7
Aug. 7, 1983	166	12.0	12.0	11.3	11.3	11.8

1 - Decay model includes Equations [4.14], [4.15], and [4.16].

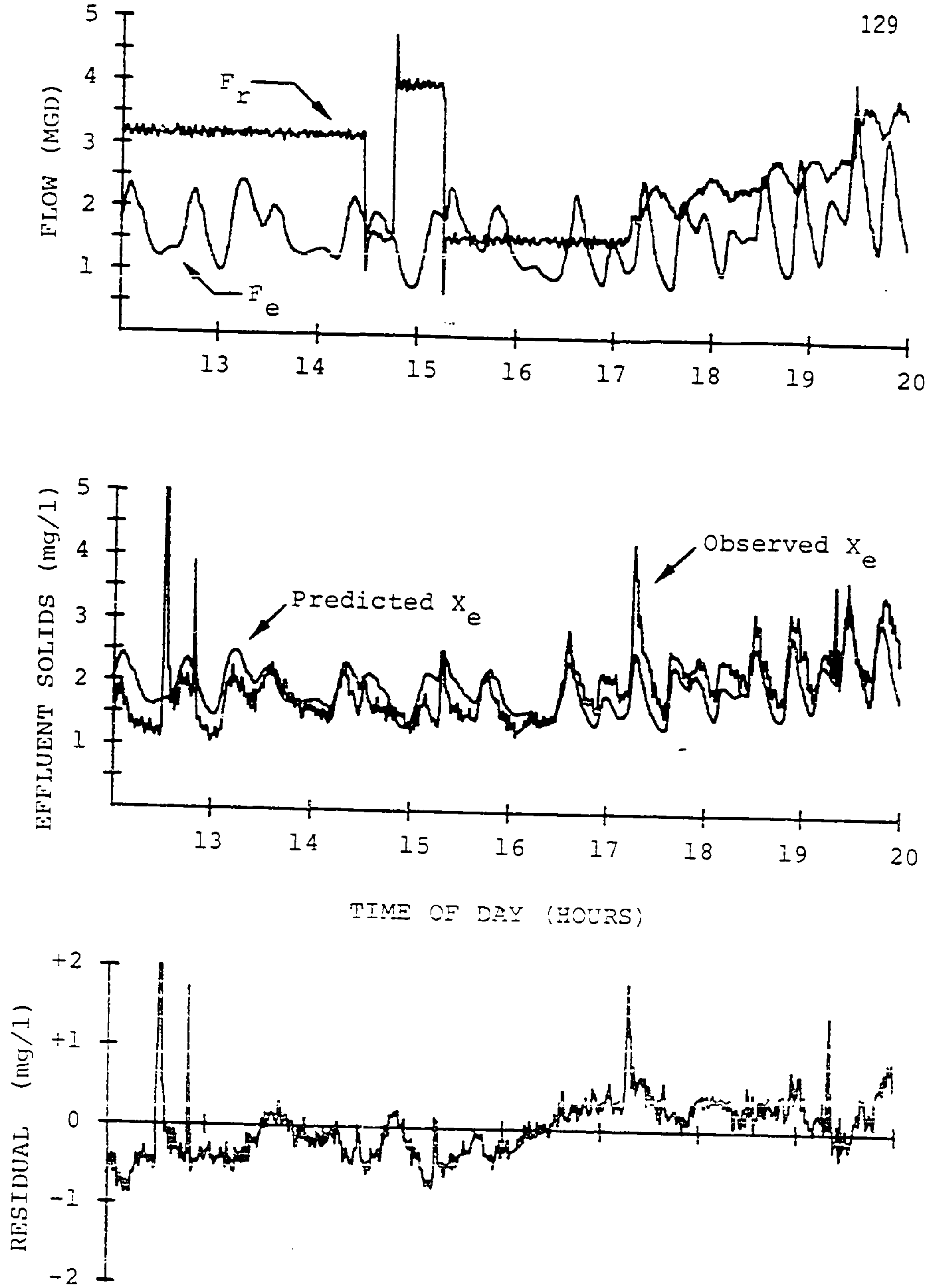

FIGURE 4.14 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATION [4.10] - APRIL 21, 1983

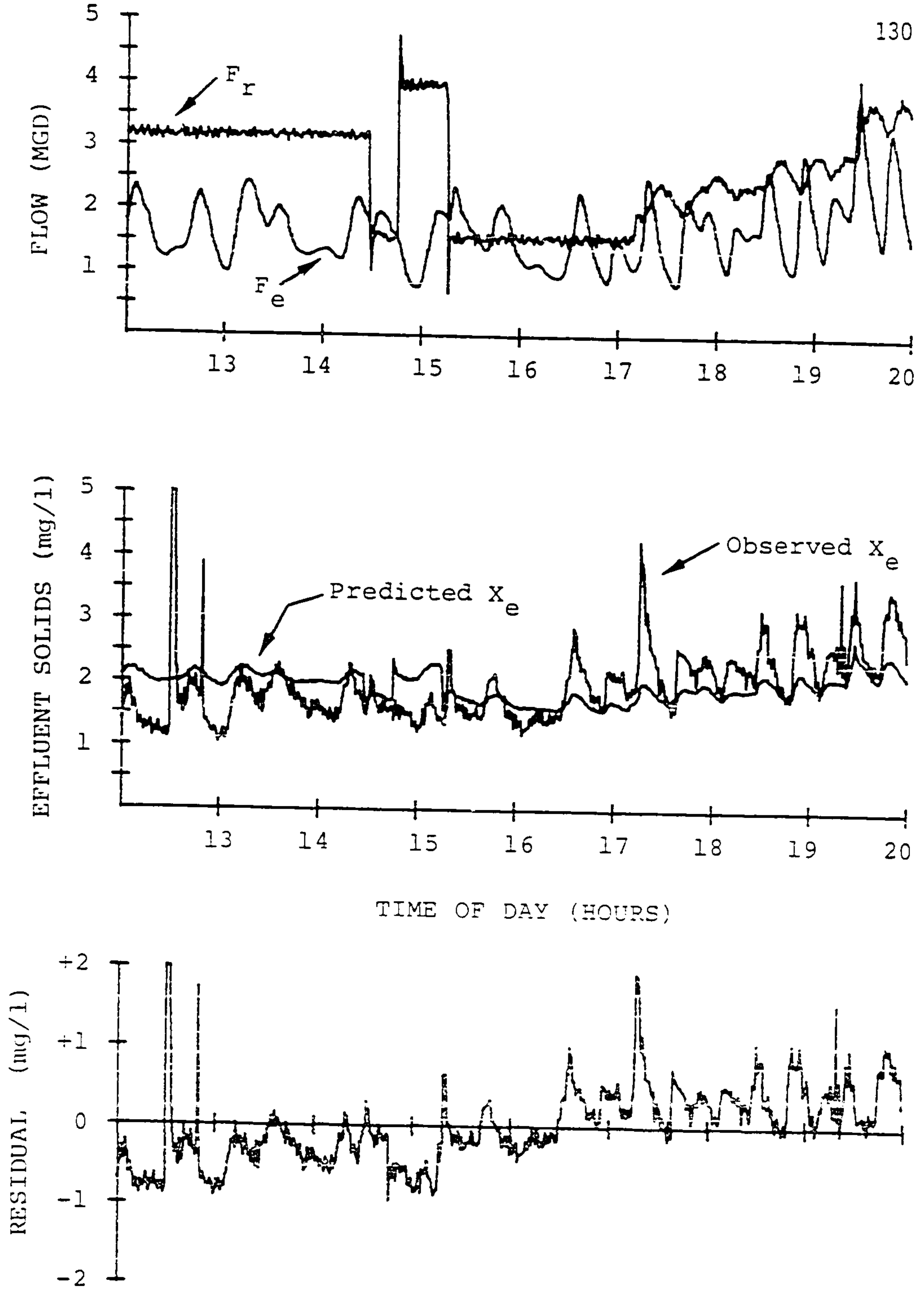

FIGURE 4.15 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATION [4.11] - APRIL 21, 1983

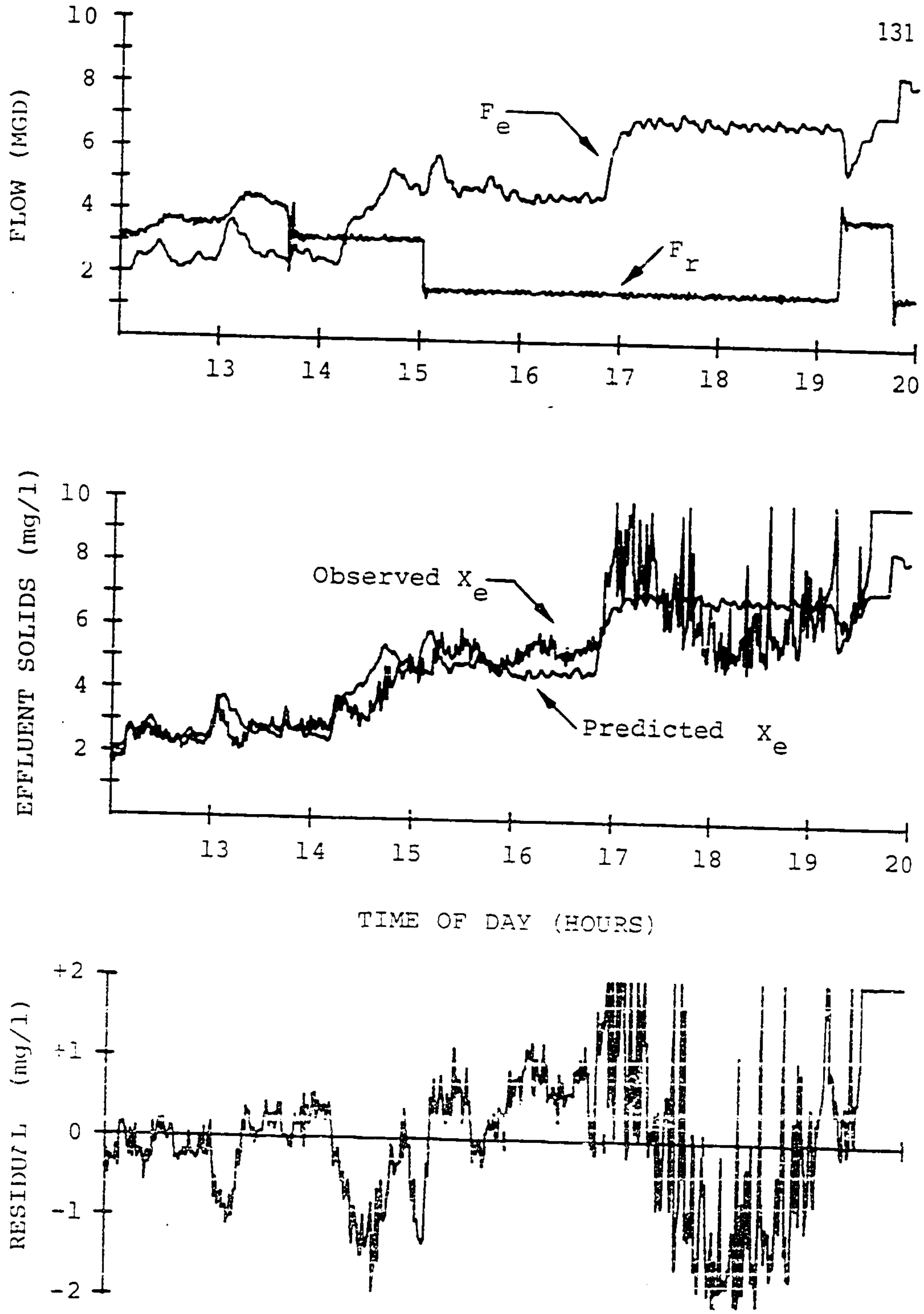

FIGURE 4.16 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATION [4.10] - APRIL 23, 1983

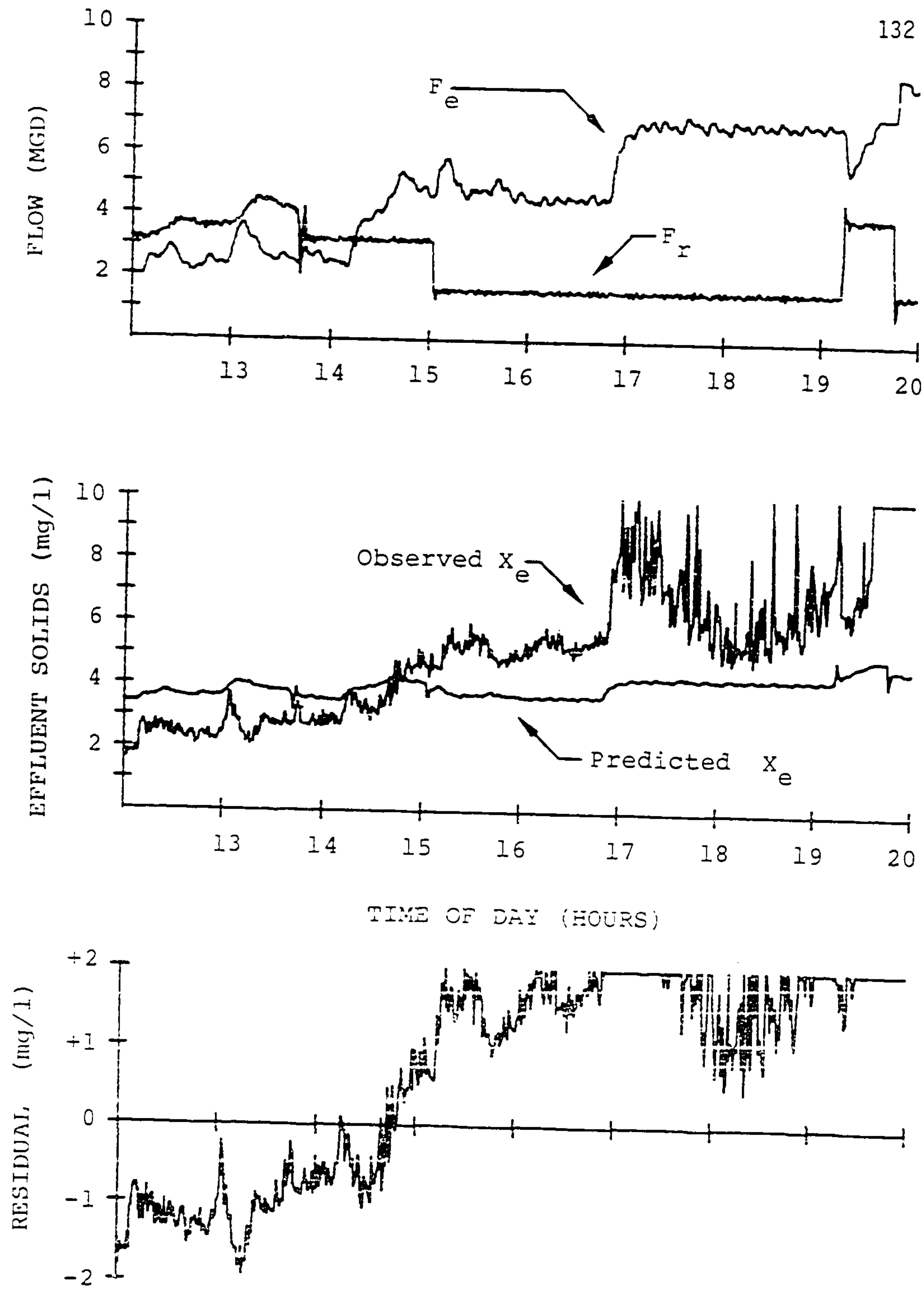

FIGURE 4.17 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATION [4.11] - APRIL 23, 1983

of effluent quality due to increased turbulance in the settler. It was not possible to conduct tests in this range at the Sagemont facility.

Effects of MLSS Concentration

It was noted that there is controversy over the effects of MLSS concentration on effluent solids. Pflanz (95), Ghobrial (54), and Chapman (20) found that higher MLSS concentrations increased Xe while Agnew (1) found that it decreased Xe. Two additional models were evaluated to quantify the short term (hours to days) effects of MLSS concentration on Xe. They are:

$$Xe = K_1 + K_2 \cdot Fe \cdot MLSS \qquad [4.12]$$

$$Xe = K_1 + K_2 \cdot (Fe+Fr) \cdot MLSS \qquad [4.13]$$

Tables 4.7 and 4.8 show the least squares estimate of the parameters for Equations [4.12] and [4.13] respectively. These models were compared to Equations [4.10] and [4.11] in Table 4.6. Of the experiments without feed forward, only that of April 26, 1983 (Figure 4.8) showed a large variation of MLSS (7,500 to 4,000 mg/l). For this day, however, the inclusion of MLSS as an input did not improve the prediction of Xe. Based on this limited evidence, the MLSS concentration can be lumped with the K_2 parameter (as in Equation [4.10]). Again, the model excluding recycle flow (Eq. [4.12]) provided better prediction of effluent solids concentration.

Over the long term (many days), two phenomena were observed which related to MLSS concentration. First, the minimum Xe (obtained at

TABLE 4.7

PARAMETERS FOR EQUATION [4.12]

$$Xe = K_1 + K_2 \cdot Fe \cdot MLSS$$

Date	K_1	K_2	NUMBER DATA POINTS	SUM RESIDUAL SQUARED	AVG ERROR
April 21, 1983	0.70	0.9309E-4	480	59.9	0.35
April 23, 1983	-0.04	1.4820E-4	285	74.0	0.51
April 26, 1983	1.88	0.6334E-4	480	294.5	0.78
Aug. 2, 1983	1.84	0.4071E-4	480	29.1	0.25
Aug. 3, 1983	1.83	0.3158E-4	480	24.8	0.23
Aug. 7, 1983	1.72	0.4618E-4	166	11.3	0.26

TABLE 4.8

PARAMETERS FOR EQUATION [4.13]

$$Xe = K_1 + K_2 \cdot (Fe+Fr) \cdot MLSS$$

Date	K_1	K_2	NUMBER DATA POINTS	SUM RESIDUAL SQUARED	AVG ERROR
April 21, 1983	0.80	0.3401E-4	480	102.7	0.46
April 23, 1983	1.82	0.4297E-4	285	424.7	1.22
April 26, 1983	1.07	0.5815E-4	480	405.4	0.92
Aug. 2, 1983	1.75	0.2833E-4	480	39.9	0.29
Aug. 3, 1983	1.64	0.2960E-4	480	23.4	0.22
Aug. 7, 1983	1.29	0.4833E-4	166	11.3	0.26

Fe=0.0) does not appear to be greatly affected my MLSS concentration. Observed minimum effluent solids were around 2.0 mg/l during the April tests when MLSS was 7,000-8,000 mg/l and about 1.5 mg/l during the August tests when MLSS was 3,000-4,000 mg/l. Secondly, the highest Xe for similar forcings was about 7 mg/l for the April tests and 3.5 for the August tests. Thus it appears that the flow proportional variable (K_2 in Equation [4.10]) is roughly proportional to MLSS concentration.

In summary, over long time spans (weeks) effluent suspended solids concentration changes approximately proportional to MLSS. This may be handled by including a MLSS multiplier in the equation (Eq. [4.12]) or by dynamically adjusting the parameter K_2 of the simpler model (Eq. [4.10]). A technique for recursively estimating K_2 is presented in a later section.

Effects of Feed Forward Flow

Two experiments were conducted using the feed forward capabilities at the Sagemont facility. These experiments are summarized in Figures 4.9 (June 25, 1983) and Figure 4.10 (July 27, 1983). In both tests a significant amount of the raw sewage flow was pumped directly to the third reactor. In each experiment the feed forward flow had predictable results on the MLSS concentrations. However, higher than normal values of effluent suspended solids (increases from 3.5 to 8.0 mg/l) were recorded during the tests. Yust and Murphy (143) noted a similar effect but contributed it to over-aeration in their pilot plant.

The exact reason for this increase in effluent solids is not known. However, Starkey (108) noted similar effects due to low D.O. concentrations. He hypothesized that additional effluent solids were primary particles present in the influent sewage. Under low D.O. concentrations the sorption of these particles was inhibited. The present research at least partially supports Starkey's hypothesis. During the June test, the D.O. concentrations in reactors 3 and 4 decreased to near zero while D.O. was maintained greater than 2.5 mg/l in the July test. The June test showed a much greater increase in effluent solids (3.5 to 8.0 mg/l) than the July test (3.0 to 5.0).

This phenomenon of increased effluent solids due to the use of feed forward flow was unexpected and outside the scope of this investigation. No attempt was made to model this mechanism. This phenomenon does, however, raise the question as to how fast and under what conditions influent particulates are actually adsorbed. Clifft (23) found particulate adsorption to be very fast while working with low sludge age solids in a high D.O. environment (UNOX process). Perhaps under other conditions, this adsorption rate would be different or desorption would occur.

Effects of Sludge Blanket Height

Ghobrial (54) is the only investigator to specifically model the effects of sludge blanket level on effluent solids. He found that if the sludge rose above a critical level, effluent quality deteriorated. In order to test this hypothesis at full scale, two flow pulses were applied to the settler. Between the two pulses, the sludge blanket was allowed to build up. Thus, one pulse was applied with a low sludge blanket and the other with a high blanket.

Figure 4.18 shows the data taken during this test including effluent flow rate, effluent suspended solids concentration, and sludge blanket level. According to Ghobrial's model, the effluent solids during the second pulse should be slightly more than one and one half times that of the first pulse. However, the response to both forcings are, in fact, very similar. In other words, the height of the sludge blanket (over the range studied) did not enhance or decrease solids capture.

This apparent discrepency between the present research and published literature can be partially explained by examining the conditions under which each experiment was conducted. The present research was conducted at an operating WWTF with a 100 foot (30 m) diameter clarifier. In the experiment reported, the sludge blanket rose to within 4.3 feet (1.3 M) of the water surface with an overflow rate of 1100 gpd/sq ft (1.9 m/hr) and a solids loading of 70 lbs/sq ft/day (340 kg/sq m/day). All inputs and outputs were continuously monitored with on-line instrumentation with frequent manual samples taken for checking the instrument calibration.

Ghobrial, on the other hand, did his work at bench scale using sludge from an operating plant. This sludge, however, was aerated for 24 hours before use. Aerating at bench scale typically results in very high shear stresses and fragmented flocs. Ghobrial also introduced the inflow near the bottom of his settling apparatus in an upward direction within the sludge blanket. This undoubtedly caused more turbulance in

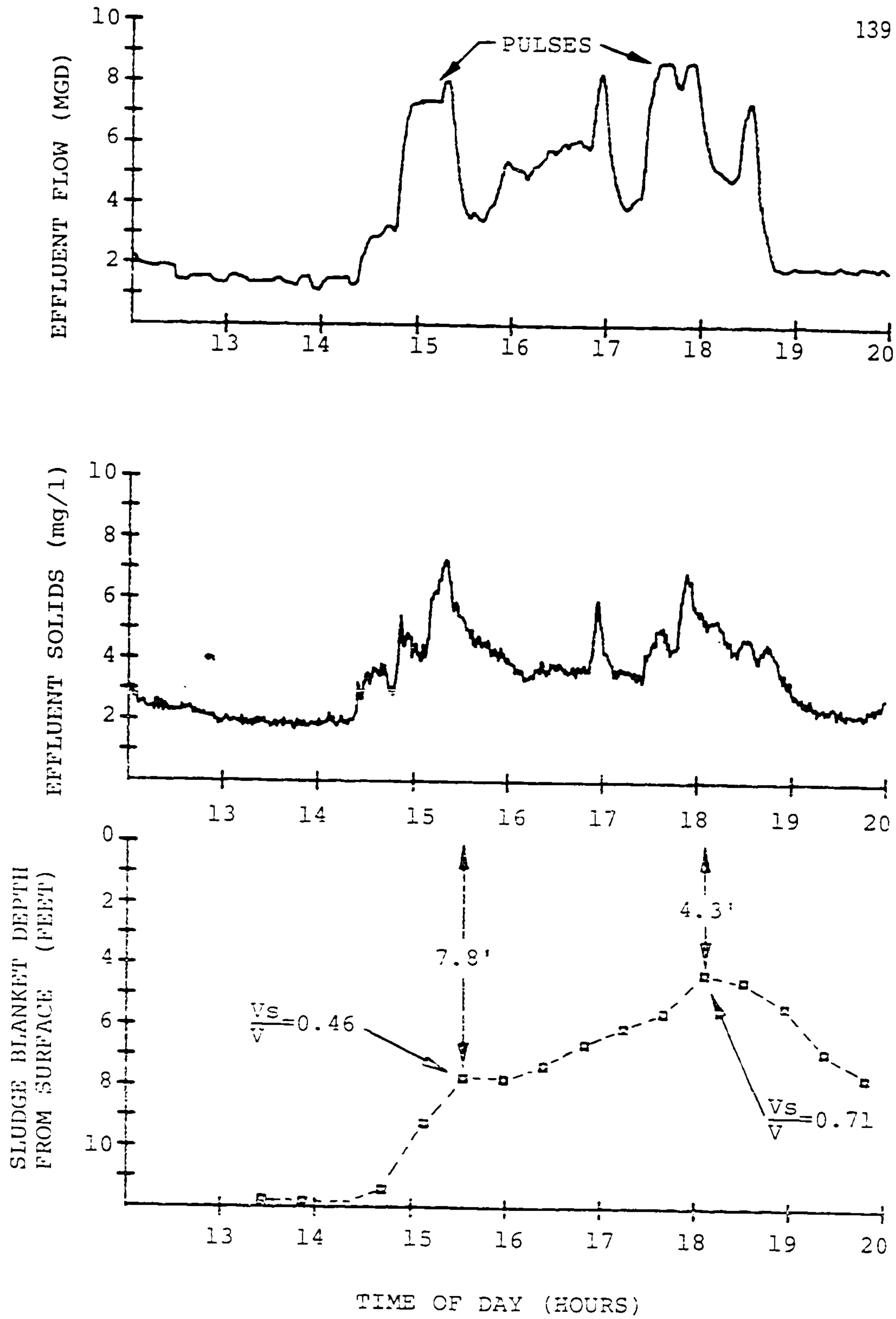

FIGURE 4.18 EFFECT OF SLUDGE BLANKET HEIGHT ON EFFLUENT SUSPENDED SOLIDS - APRIL 26, 1983

the sludge blanket than would be encountered in practice. Perhaps these two factors also explain the unusually high solids levels (up to 290 mg/l) reported. Also, the settling apparatus Ghobrial used was only 14 inches high (35 cm) and 3.0 inches (7.5 cm) in diameter.

Effects of Hydraulic Transients

Several previous researchers (26, 87) have noted that hydraulic transients have an immediate effect which tends to die out slowly. The present research supports this concept. Sudden increases in flow result in an immediate increase in effluent solids. Sudden decreases in flow, however, are accompanied by a slower decay of effluent solids.

An improvements to Equation [4.10] was sought to give this type of behavior. For increasing or slowly decreasing flow, it is assumed that Equation [4.10] holds.

$$(Xe)_{ss} = K_1 + K_2 \cdot Fe \qquad [4.14]$$

where:

$(Xe)_{ss}$ = Xe at steady state conditions (M/L^3).

However, if the value predicted in Equation [4.14] is less than the present effluent solids concentration (as with a sudden decrease in flow), the decay of Xe is assumed to be proportional to the difference between the present solids concentration and the steady state value.

$$\text{if } (Xe)_{ss} > Xe \text{ then } Xe = (Xe)_{ss} \qquad [4.15]$$

$$\text{if } (Xe)_{ss} < Xe \text{ then } \frac{d(Xe)}{dt} = K_3 \cdot [Xe-(Xe)_{ss}] \qquad [4.16]$$

where:

K_3 = decay rate coefficient (1/T).

Since Equation [4.16] applies only when (Xe)ss is less than Xe, the parameter K_3 should have a negative value. Collectively, Equations [4.14], [4.15], and [4.16] are hereafter referred to as the decay model.

Table 4.9 shows the parameters derived for this model. It should be noted that the data from April 23 and August 3, 1983 show decay constants significantly different from the other experiments. On April 23, 1983 the flow was steadily increased for the duration of the experiment. On August 3, 1983 the flow was varied at a slow rate. Thus it is not surprising that the data from these days do not yield a good estimate for a parameter associated with sudden decreases in flow. The positive value of K_3 on April 23, 1983 is especially misleading since it signifies that Xe increases with a decreasing flow rate. This would result in an unstable process and is therefore not reasonable.

Plots of the model estimate based upon Equations [4.14], [4.15], and [4.16] are shown in Figures 4.19 through 4.22.

One feature not observed in the Sagemont data is that of large overshoots in Xe due to rapid increases in flow as reported by Chapman (21). Chapman hypothesized that these overshoots might be caused by the accumulation of floating solids on his pilot-scale clarifier which were "dislodged" with large flow increases. His clarifier did not include any skimmer mechanism. The present research supports his hypothesis that the phenomenon was induced by the experimental apparatus.

TABLE 4.9

DECAY MODEL

PARAMETERS FOR EQUATIONS [4.14], [4.15], and [4.16]

DATE	K_1	K_2	K_3	AVG ERROR
April 21, 1983	0.65	0.7226	-0.2184	0.37
April 23, 1983	-3.32	0.9098	0.0035	0.36
April 26, 1983	1.34	0.4753	-0.0465	0.52
Aug. 2, 1983	1.70	0.1656	-0.1091	0.20
Aug. 3, 1983	1.80	0.1090	-5.2084	0.21
Aug. 7, 1983	1.77	0.1363	-0.2485	0.27

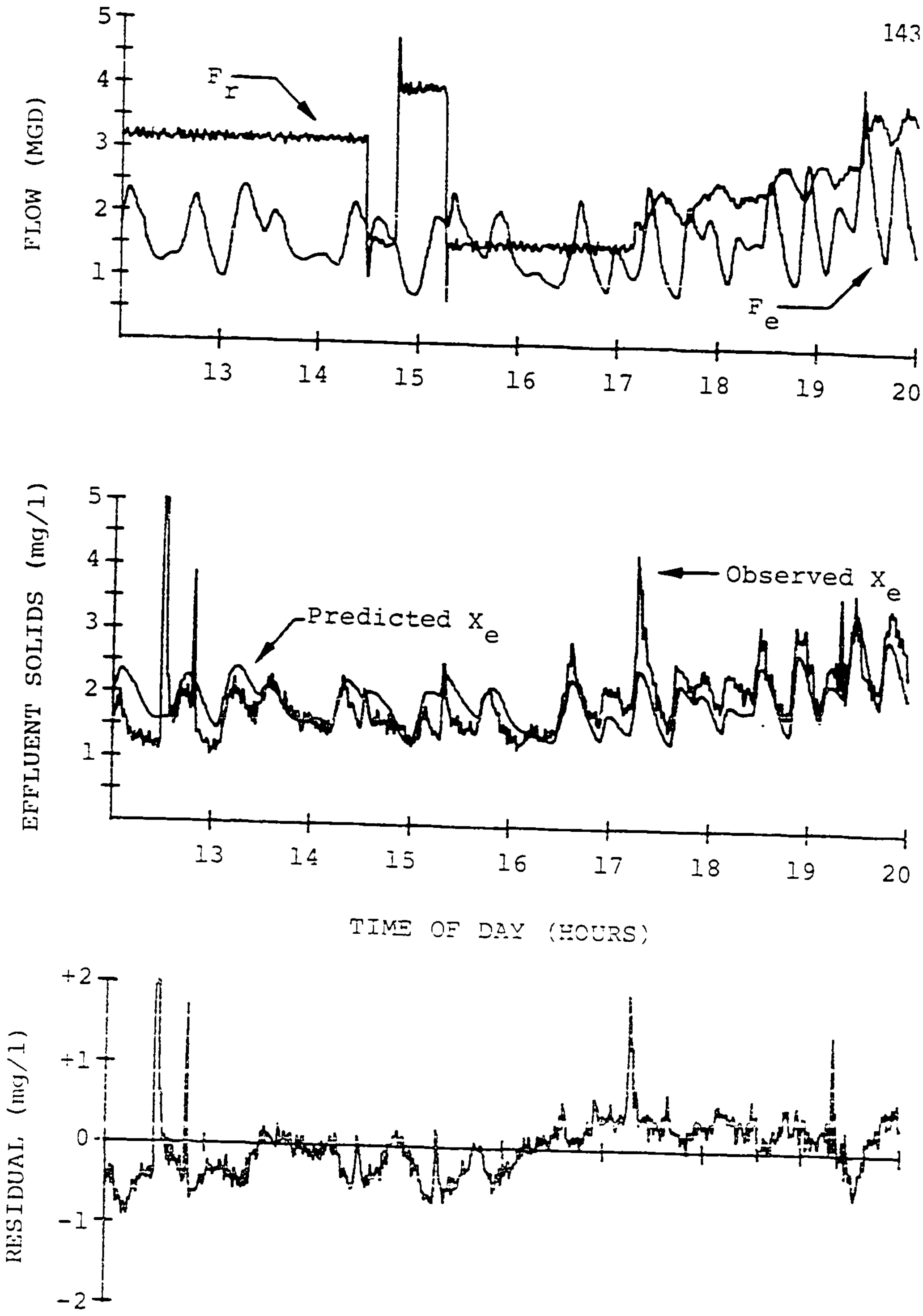

FIGURE 4.19 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON THE DECAY MODEL - APRIL 21, 1983

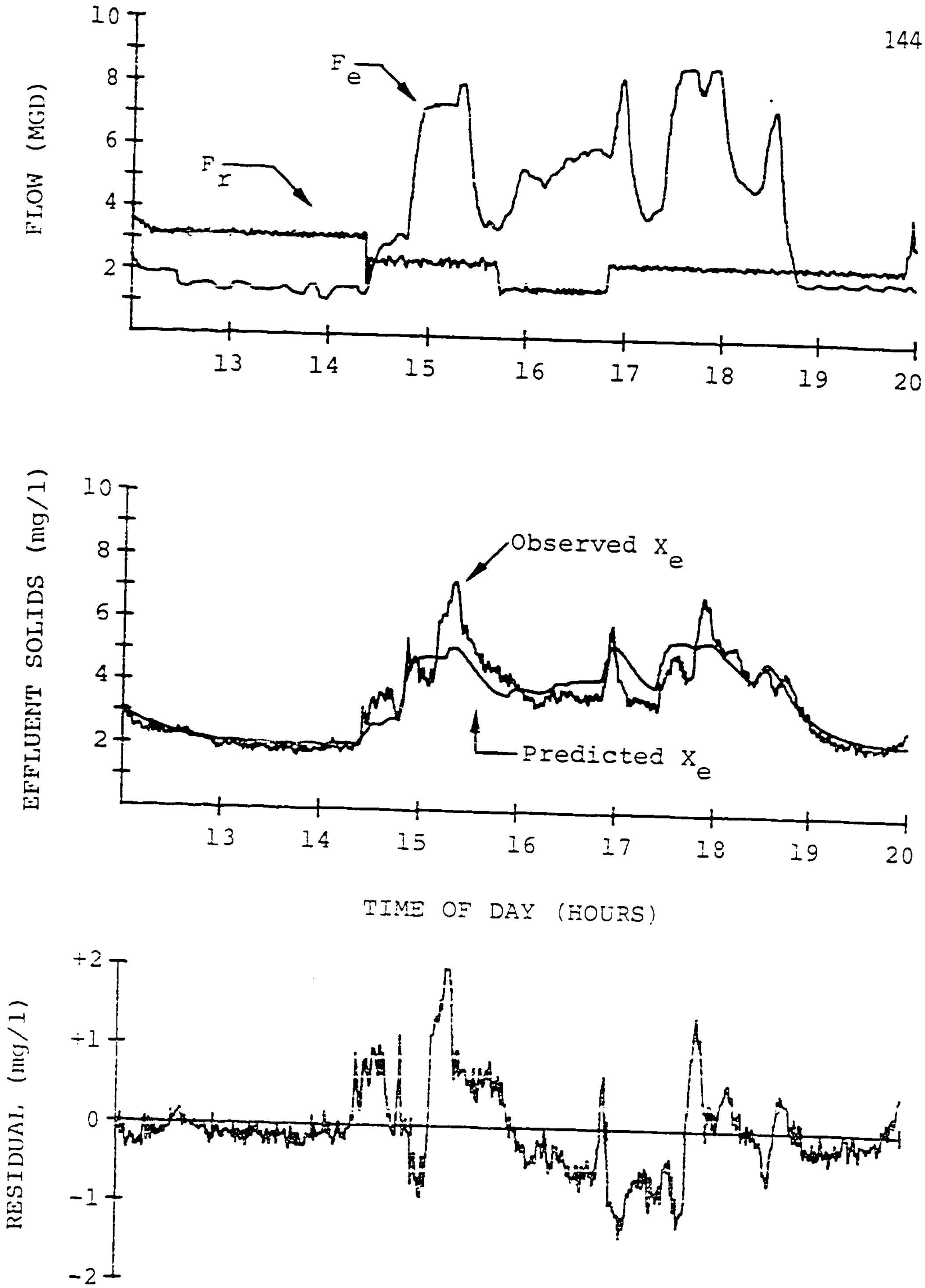

FIGURE 4.20 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON THE DECAY MODEL - APRIL 26, 1983

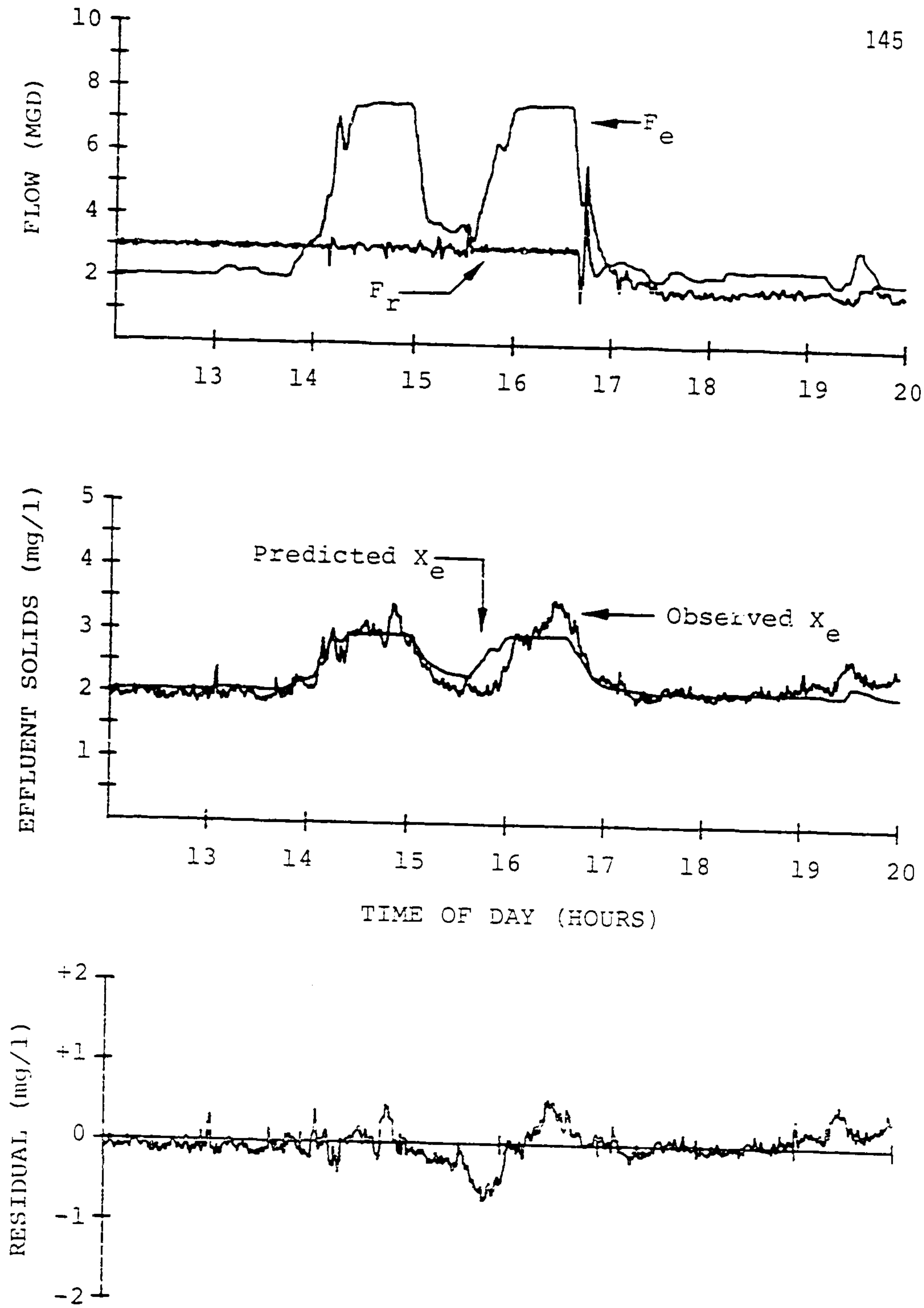

FIGURE 4.21 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON THE DECAY MODEL - AUGUST 2, 1983

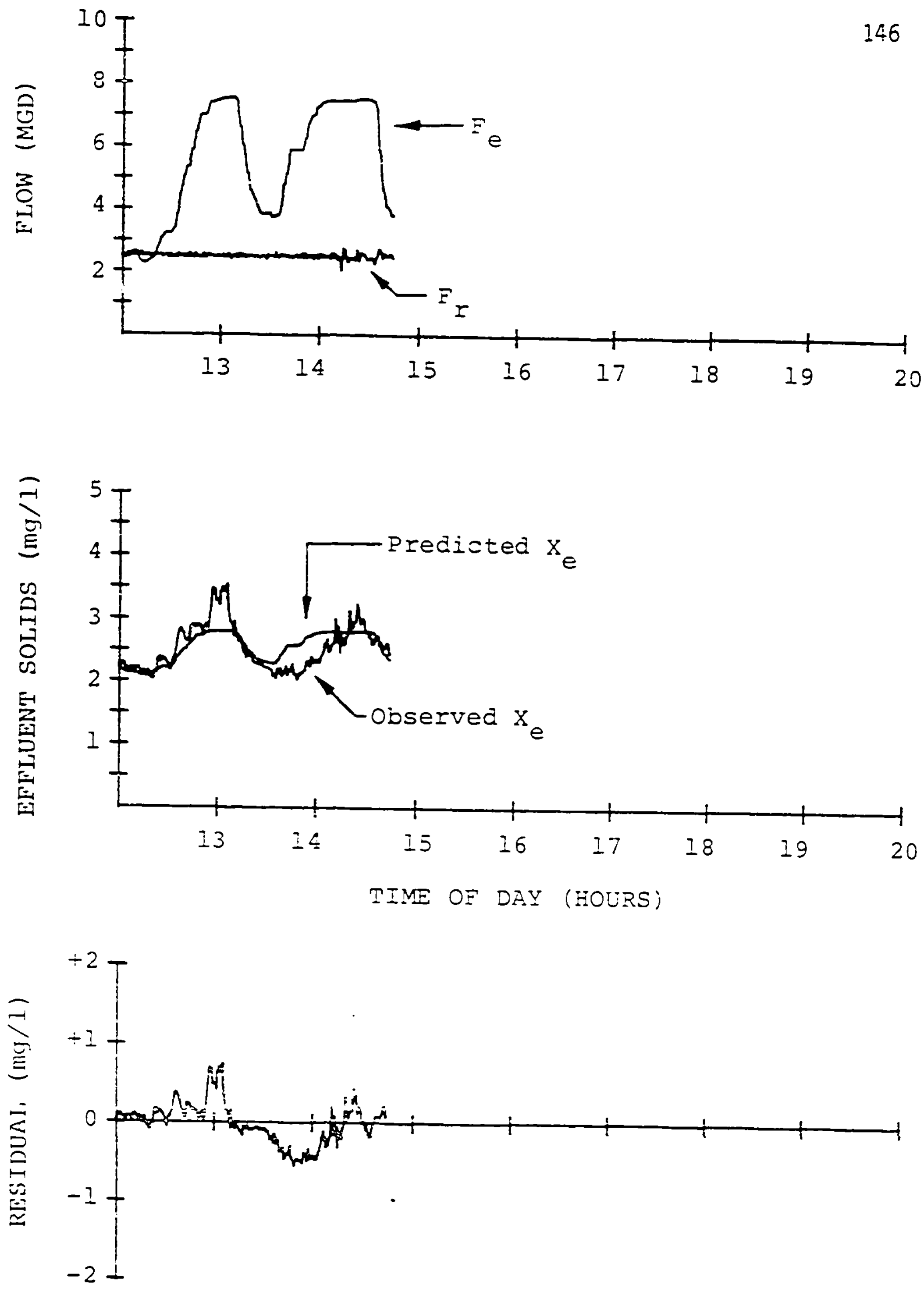

FIGURE 4.22 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON THE DECAY MODEL - AUGUST 7, 1983

Clarification Model Summary

The decay model as defined by Equations [4.14], [4.15], and [4.16] is proposed as the clarification model for further work herein. This model is capable of simulating many of the process features observed at the Sagemont plant such as the rapid increase and slow decay of Xe with flow transients. While it is possible that additional improvements to the model could be made, it is felt that the proposed model is adequate for the purposes of this study and is preferred over the other possible models because of its simplicity. A number of other models (see Table 4.10) including time lags, combinations of parameters, and first and second order terms were also evaluated in this investigation but were rejected because they gave no significant improvement over the decay model.

Thickening Model

Stenstrom's (110) thickening model was chosen for implementation for several reasons. Of the derivatives from Bryant's work (16, 110, 121), Stenstrom's model gave a better description of the observed data. Unlike George's (53) model, Stenstrom's is capable of predicting a rising sludge blanket without any artificial constraints. Also, both George's and Kos' (71) models were considered too complicated for implementation on a small process control computer. It is possible to implement Stenstrom's model on a small computer since only 10 to 20 ordinary differential equations must be numerically solved. The adequacy of the model was evaluated by comparing simulated results with experimental data.

TABLE 4.10

ADDITIONAL CLARIFICATION MODELS
WHICH WERE EVALUATED IN THIS INVESTIGATION

$$Xe = K_1 + K_2 \cdot Fe + K_3 \cdot MLSS$$

$$Xe = K_1 + K_2 \cdot Fe + K_3 \cdot MLSS + K_4 \cdot Fe \cdot MLSS$$

$$Xe = K_1 + K_2 \cdot (Fe+Fr) + K_3 \cdot (Fe+Fr) \cdot MLSS$$

$$Xe = K_1 + K_2 \cdot Fe + K_3 \cdot \frac{d(Fe)}{dt}$$

$$\frac{d(Xe)}{dt} = K_1 \cdot Xe + K_2 \cdot Fe + K_3 \cdot \frac{d(Fe)}{dt}$$

$$\frac{d(Xe)}{dt} = K_1 \cdot Fe \cdot Xe + K_2 \cdot Fe \cdot MLSS + K_3 \cdot \frac{d(Fe)}{dt}$$

$$\frac{d(Xe)}{dt} = \frac{(Fe+Fr)}{K_3} \cdot (K_1 + K_2 \cdot Fe - Xe)$$

$$\frac{d(Xe)}{dt} = \frac{(Fe)}{K_3} \cdot (K_1 + K_2 \cdot Fe - Xe)$$

$$\frac{d^2(Xe)}{dt^2} = K_1 \cdot \frac{d(Xe)}{dt} + K_2 \cdot Xe + K_3 \cdot Fe + K_4 \cdot \frac{d(Fe)}{dt}$$

Development of Thickening Model

Stenstrom's thickening model was reviewed in section II along with the six assumptions on which the model is based. Modifications to this model will be presented in a following section.

The model (Equation [2.18]) consists of a number of simultaneous, ordinary differential equations with boundary conditions imposed on the top and bottom elements. The equation for the top finite volume becomes:

$$\frac{d(X_1)}{dt} = \frac{FLUXIN - U \cdot X_1 - MIN(Gs_1, Gs_2)}{Z} \qquad [4.17]$$

where:

X_1 = solids concentration in top element (M/L^3),

FLUXIN = net solids flux density into settler ($M/T/L^2$),

U = bulk downward velocity (L/T),

MIN = minimum function,

Gs = settling flux, $Vs \cdot X$ ($M/T/L^2$),

Z = finite depth of settler element (L).

The net flux entering the thickening zone of the settler is described as Equation [4.18]. For most practical situations the Fe•Xe term can be neglected with less than one percent error.

$$FLUXIN = [(Fe+Fr) \cdot MLSS - Fe \cdot Xe]/AREA \qquad [4.18]$$

For each finite volume of the settler except the top and bottom elements, the equation is:

$$\frac{d(X_i)}{dt} = \frac{U\cdot(X_{i-1} - X_i)}{Z} + \frac{MIN(Gs_{i-1},Gs_i) - MIN(Gs_i,Gs_{i+1})}{Z} \qquad [4.19]$$

where:

i = subscript for any element except top or bottom.

Since the settling flux at the bottom of the settler is assumed to be zero, Equation [4.19] simplifies to:

$$\frac{d(X_n)}{dt} = \frac{U\cdot(X_{n-1}-X_n) + Gs_n}{Z} \qquad [4.20]$$

where:

n = subscript for bottom element.

The function relating solids concentration and settling velocity is assumed to have an exponential form.

$$Vs = V_0 \exp(-b\cdot X) \qquad [4.21]$$

where:

V_0, b = settling parameters.

Settling Tests

Two sets of settling tests were run; one for the April series of tests and one for the August series of tests. These settling tests were conducted in a 3-liter SSVI apparatus as described in section III. The results are shown in Tables 4.11 and 4.12 along with the least squares estimate of the settling parameters.

TABLE 4.11

SETTLING TESTS (APRIL 27, 1983)

MLSS CONCENTRATION (mg/l)	INITIAL SETTLING VELOCITY (m/hr)
1,902	5.33
2,754	4.01
3,663	2.71
4,840	1.95
5,750	1.41
6,730	1.04
8,150	0.69

V0 = 9.61
b = 3.2E-4

TABLE 4.12

SETTLING TESTS (JULY 26, 1983)

MLSS CONCENTRATION (mg/l)	INITIAL SETTLING VELOCITY (m/hr)
3,645	2.59
5,880	1.22
8,370	0.64

V0 = 7.36
b = 2.95E-4

Simulation Results

Simulations were conducted using the settler model (with 19 elements), the settling parameters from that period, the measured MLSS concentrations, and the measured effluent and recycle flows. The underflow concentrations from these simulations and the measured return activated sludge (RAS) concentrations are shown as Figures 4.23 through 4.26. The simulations were started four hours before the start of the plots in order to minimize the effects of initial conditions.

Figures 4.23 and 4.24 generally show good correlation between the measured and predicted values. Several deficiencies, however, are apparent. The model is not capable of predicting the RAS discontinuities associated with the large step changes in the recycle flow rate. Like many models, this one has problems with discontinuities.

The experimental return sludge concentrations show periodic peaks on a very regular basis (12-5/6 minutes) as shown in Figure 4.27. The time between peaks corresponds to one half the revolution time (25-2/3 minutes) of the sludge rake. It is hypothesized that heavy solids are sweep into the sludge hopper every time an arm of the sludge rake passes. This temporarily increases the solids concentration. For conditions in which a sludge blanket exists, this swing in Xr may be as little as 500 to 1000 mg/l. When no sludge blanket exists, the swing may be as much as 5,000 to 6,000 mg/l. A similar phenomenon was recently observed by Chapman (21) at pilot scale. To the author's knowledge, however, this is the first documented accounting of this

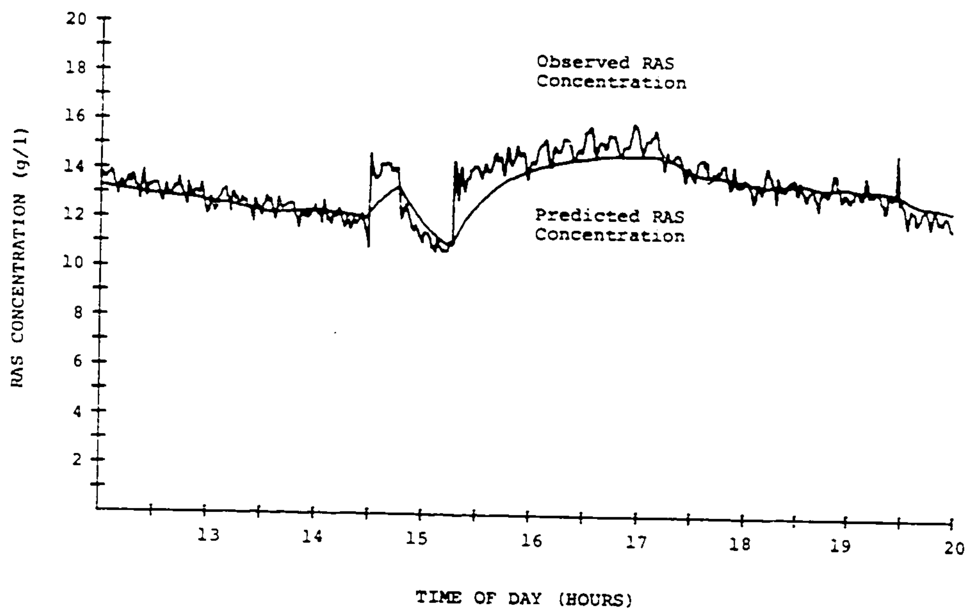

FIGURE 4.23 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED UPON STENSTROM'S MODEL - APRIL 21, 1983

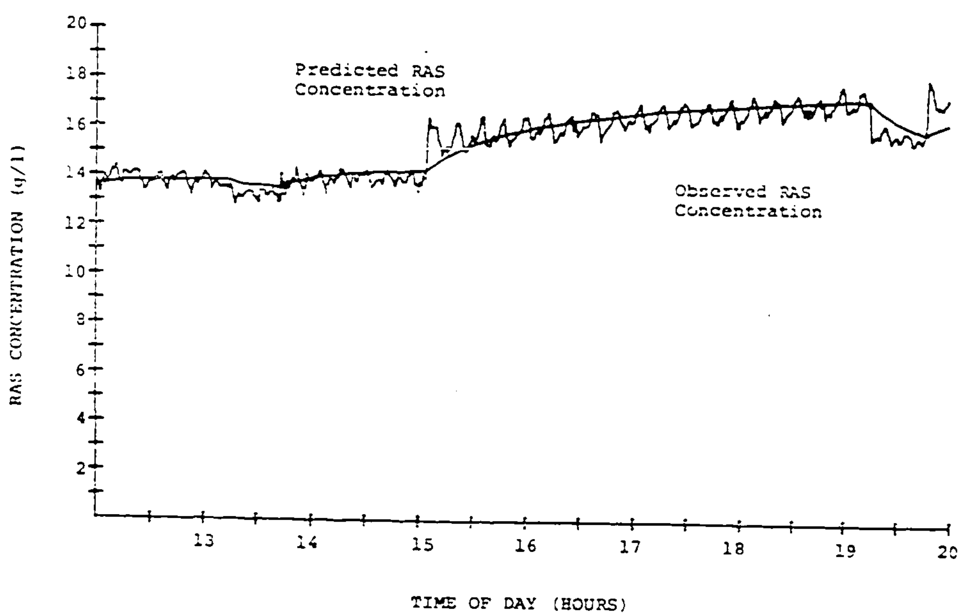

FIGURE 4.24 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED UPON STENSTROM'S MODEL - APRIL 22, 1983

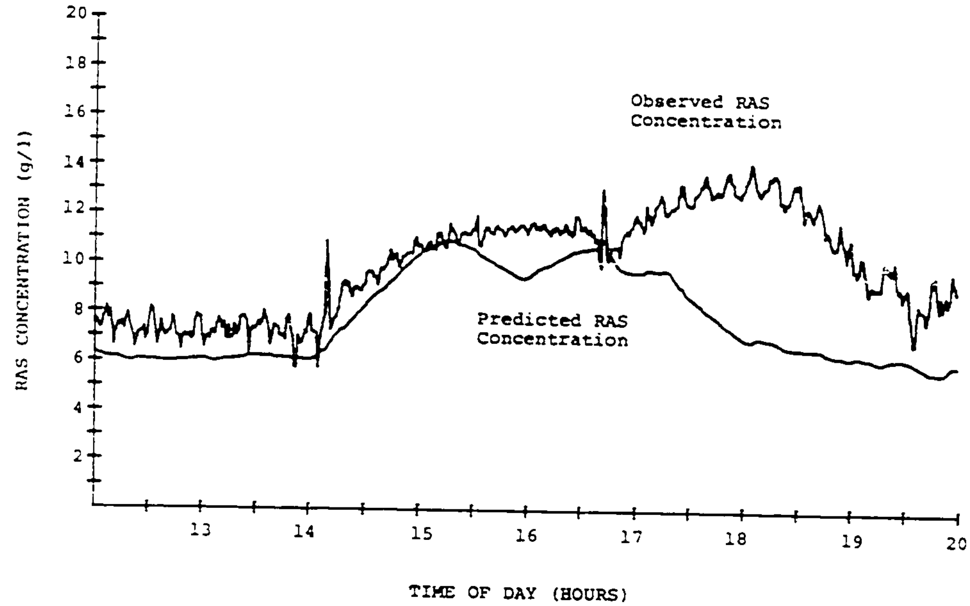

FIGURE 4.25 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED UPON STENSTROM'S MODEL - AUGUST 2, 1983

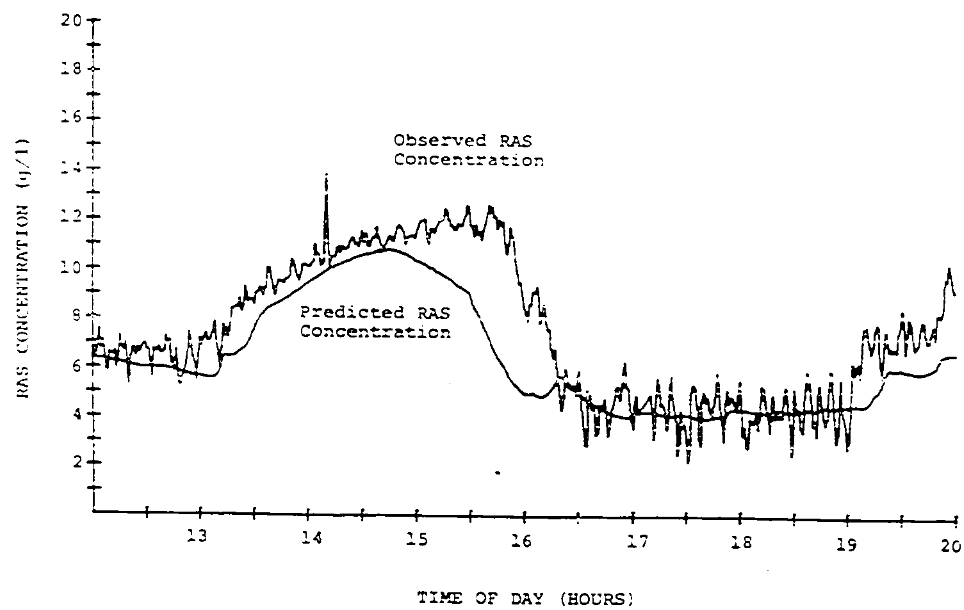

FIGURE 4.26 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED UPON STENSTROM'S MODEL - AUGUST 7, 1983

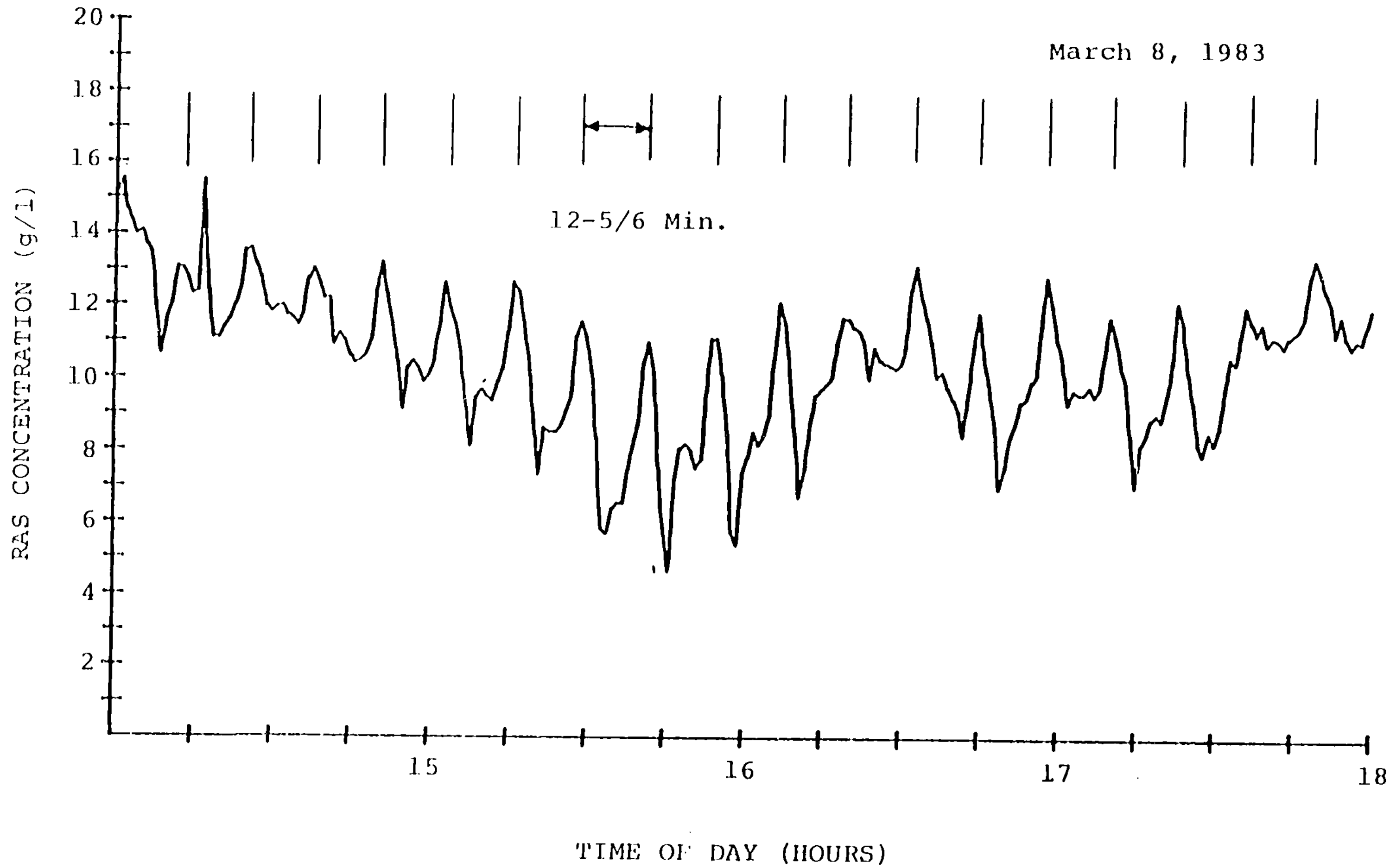

FIGURE 4 27 PERIOD PEAKS IN RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION phenomenon at a full-scale wastewater treatment plant. The model includes no mechanism to simulate these short term fluctuations. This phenomenon and its effects on sampling and control will be discussed in a later section.

Figures 4.25 through 4.27 also show an approximate correlation between measured and predicted underflow concentrations. However, it is noted that the predicted underflow concentrations are generally less than the measured ones. This discrepency led the author to believe there could be some error in one or more of the measurements. A 24-hour mass balance around the settler indeed showed this. For the August tests it was found that the output solids (effluent and underflow) exceeded the input by up to 17 percent. This error greatly exceeded the 4 percent maximum error encountered during the April tests. It is suspected that the measurement error was associated with one of the flow measurements (effluent or recycle) since the solids monitors were manually checked for accuracy before and during the testing.

Thickening Model Modifications

Even after taking the possible measurement errors into account, several of the experiments still show discrepancies between the predictions and measurements. Figure 4.26 (August 2, 1983) especially shows the model and process measurements diverging in different directions. This discrepency is thought to be caused, at least in part, by the conical section of the settler. This section accounts for 37 percent of the settler height and 17 percent of the volume. As the sludge travels down the conical section, the cross-sectional area changes nonlinearly.

Therefore the relative importance of sludge settling and bulk transport also change in a nonlinear manner relative to the height in the cone.

The proposed modification is an extension of Stenstrom's model using the same assumptions to develop equations for the conical section of the settler. Stenstrom's equations ([4.17] and [4.19]) are used for the model elements in the cylindrical section of the settler. The new model equations will be used in the conical section.

Figure 4.28 shows a differential volume of the conical section of the settler. A mass balance around the "ith" element results in the following:

$$\Delta V_i \cdot \frac{d(X_i)}{dt} = \Delta A_{i-1} \cdot (Vs_{i-1} + U_{i-1}) \cdot X_{i-1} - \Delta A_i \cdot (Vs_i + U_i) \cdot X_i \qquad [4.22]$$

where:

ΔV = volume of frustrum (L^3),

ΔA = area (L^2).

Equation [4.22] can be rearranged and simplified to give:

$$\frac{d(X_i)}{dt} = \frac{\Delta A_{i-1} \cdot U_{i-1} \cdot X_{i-1} - \Delta A_i \cdot U_i \cdot X_i}{\Delta V_i} +$$

$$\frac{\Delta A_{i-1} \cdot Vs_{i-1} \cdot X_{i-1} - \Delta A_i \cdot Vs_i \cdot X_i}{\Delta V_i} \qquad [4.23]$$

By noting that $U_i = Fr/A_i$ and applying the flux constraint, Equation [4.24] is the equivalent to Equation [4.19] for a conical section.

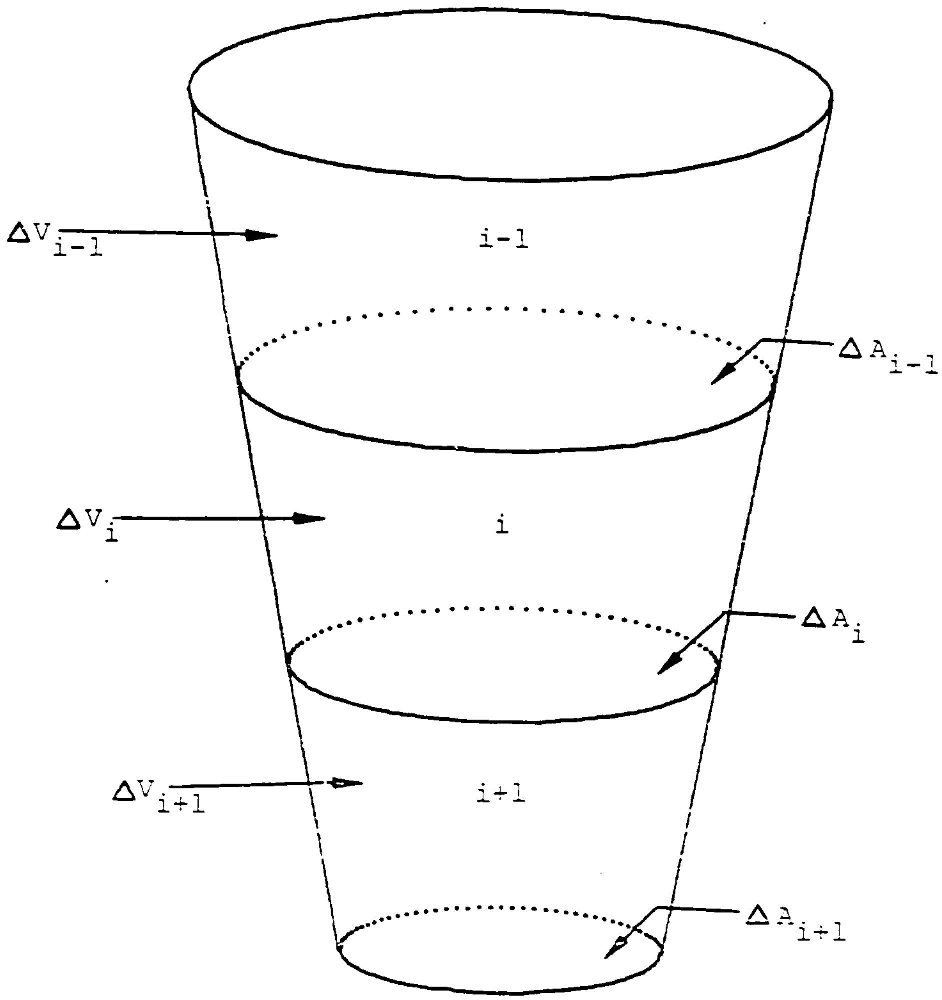

FIGURE 4.28 CONICAL DIFFERENTIAL VOLUMES

$$\frac{d(X_i)}{dt} = \frac{Fr \cdot (X_{i-1} - X_i)}{\Delta V_i} +$$

$$\frac{\Delta A_{i-1} \cdot MIN(Gs_{i-1}, Gs_i) - \Delta A_i \cdot MIN(Gs_i, Gs_{i+1})}{\Delta V_i} \qquad [4.24]$$

The equation for the bottom element now takes the form of:

$$\frac{d(X_n)}{dt} = \frac{Fr \cdot (X_{n-1} - X_n)}{\Delta V_n} + \frac{\Delta A_{n-1} \cdot Gs_n}{\Delta V_n} \qquad [4.25]$$

Simulation Results of Modified Model

The simulation results for the April series of tests are similar to those already presented as Figures 4.23 and 4.24 and are therefore not repeated. The results for the August tests, however, show considerable improvement and are shown as Figures 4.29 and 4.30. Figure 4.29 (compare with Figure 4.25), in particular, shows the simulation results more closely following the process measurement.

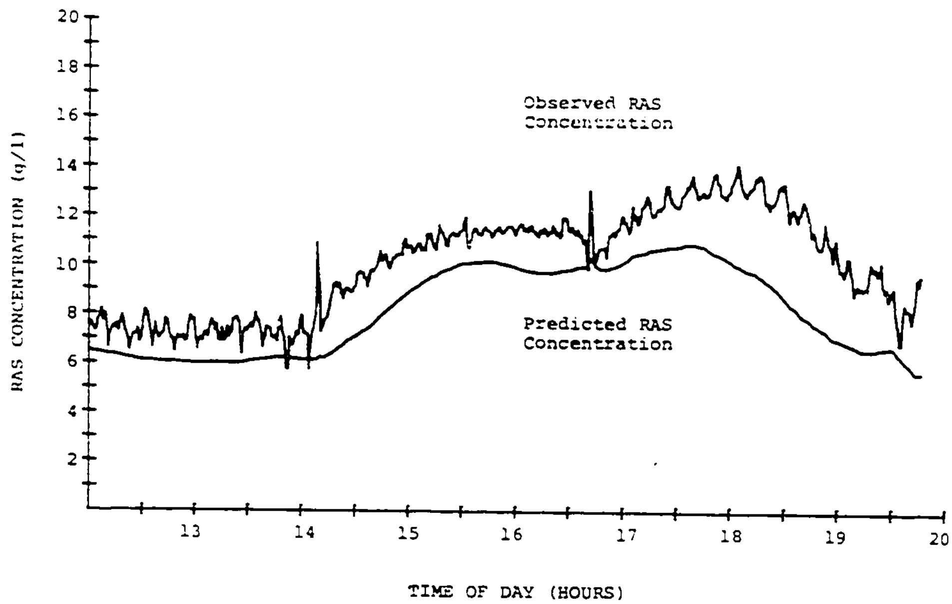

FIGURE 4.29 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED ON MODIFIED THICKENER MODEL - AUGUST 2, 1983

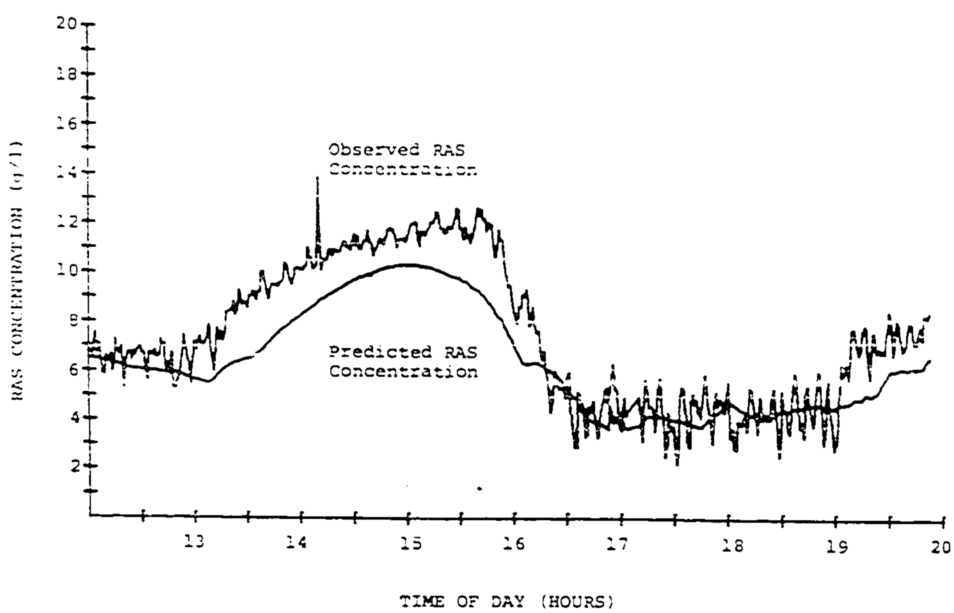

FIGURE 4.30 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION PREDICTIONS BASED ON MODIFIED THICKENER MODEL - AUGUST 7, 1983

V. PROCESS CONTROL STRATEGIES

One of the stated objectives of this investigation was to propose control strategies to improve the quality of wastewater treatment. Since the effluent quality in most properly operated municipal activated sludge system is almost entirely a function of discharged suspended solids, the objective function of these control strategies is to minimize suspended solids. It is the intent of this investigation to control suspended solids concentration by manipulation of process variables rather than by design changes or chemical addition.

There are a limited number of manipulatable control variables which affect the activated sludge process. For the purposes of this investigation only control of influent flow rate, recycle flow rate, waste flow, and feed forward flow rate will be considered. Air flow (dissolved oxygen) control will not be considered. It is assumed that sufficient air is always available. Since it was demonstrated that influent flow rate has the most effect on the suspended solids concentration, emphasis is placed on developing control strategies which affect this parameter.

Bernard (9) also emphasizes hydraulic control in the activated sludge process. He states that it is necessary to account for both the hydraulic and biological parameters in a plant. He gives examples of the effects of hydraulic parameters on the operation of an activated sludge treatment system and presents guidelines for controlling hydraulic conditions.

Waste Sludge Flow Control

Sludge age control is used for controlling the waste sludge flow rate at the Sagemont WWTF. Since it has been shown that the effluent suspended solids concentration is approximately proportional to the MLSS concentration, the set point for the sludge age should be as low as possible and still satisfy other system performance requirements (e.g., achieve nitrification and/or acceptable settling characteristics). At the plant, sludge is wasted directly from the aeration basins. This is the Garrett (50) method of sludge age control. For a sludge age of ten days, the waste pumps run approximately four hours per day. Since sludge is wasted directly from the aeration basins, the flow going to the settlers is correspondingly reduced during this time. This reduction in overflow rate during sludge wasting suggests that an effective method to reduce suspended solids is to waste during peak flow conditions. The timing of this can probably best be based upon historical flow data.

Feed Forward Flow Control

The use of the feed forward capability at the Sagemont plant has demonstrated its ability to rapidly (30-60 minutes) reduce the MLSS concentration in reactor 4 and the solids loading to the clarifier. Unfortunately, the feed forward experiments also demonstrated that effluent suspended solids increase to varying degrees depending on the dissolved oxygen concentrations in the reactors. For this reason, it is proposed that feed forward be used only to prevent gross process failure of the activated sludge process due to the sludge blanket going over the weirs (induced by high flows). Since feed forward flow is effective in reducing the solids loading to the clarifier, thickening failures of this kind can be effectively reduced. Also, under these circumstances, the increased effluent suspended solids due to the use of feed forward flow would be perferred to the large loss of solids associated with the loss of the sludge blanket.

Return Sludge Flow Rate Control

Over the range studied, it has been demonstrated that recycle flow rate has little effect on suspended solids concentration except for hydraulic transients induced by sudden changes. Therefore, return sludge flow rate should be controlled to satisfy other system performance requirements (e.g., thickening, minimization of solids stored in the settler, etc.).

Ratio control of the return sludge flow rate is a reasonable control strategy which tends to maintain constant MLSS and return sludge concentrations. The minimum recycle ratio can be approximated with Equation [5.1] which can be derived from a mass balance around the settler using volumetric concentrations.

$$r = V_{30}/(1-V_{30}) \qquad [5.1]$$

where:

r = recycle ratio (Fr/F_0),

V_{30} = MLSS 30-minute settled volume (as a fraction).

Influent Flow Rate Control

The clarification model developed in the investigation shows that effluent suspended solids concentrations increase rapidly with flow increases. Additionally, effluent solids can decrease slower than rapid flow changes. This fact suggests that flow equalization could be an effective method of reducing effluent suspended solids. This concept is shown in Figure 5.1 where simulation results using three different flow patterns, all of equal volume, are compared. Using the decay model (Equations [4.14], [4.15], and [4.16]). the average effluent suspended solids concentration is 10.05 mg/l for the single pulse, 6.71 mg/l for the dual pulses, and 4.5 mg/l for steady flow. In general, variations in influent flow rate result in higher average suspended solids concentrations than steady flows. The minimum suspended solids concentration for a given volume of wastewater is attained with a constant flow rate.

Figure 5.2 shows typical effluent flow rates (one minute averages) over the course of one day. This jagged flow curve results from the on/off operation of the influent pumps. From the previous argument, it can be concluded that the average suspended solids concentration with this type of flow pattern is much greater than that which could be attained from a more constant flow.

Although flow equalization has been widely used for industrial waste treatment, very few equalization basins are in use at municipal treatment plants. In part this is because of the large volumes of wastewater encountered at municipal plants, the large diurnal variations. the lack of demonstrated benefits, the additional costs of

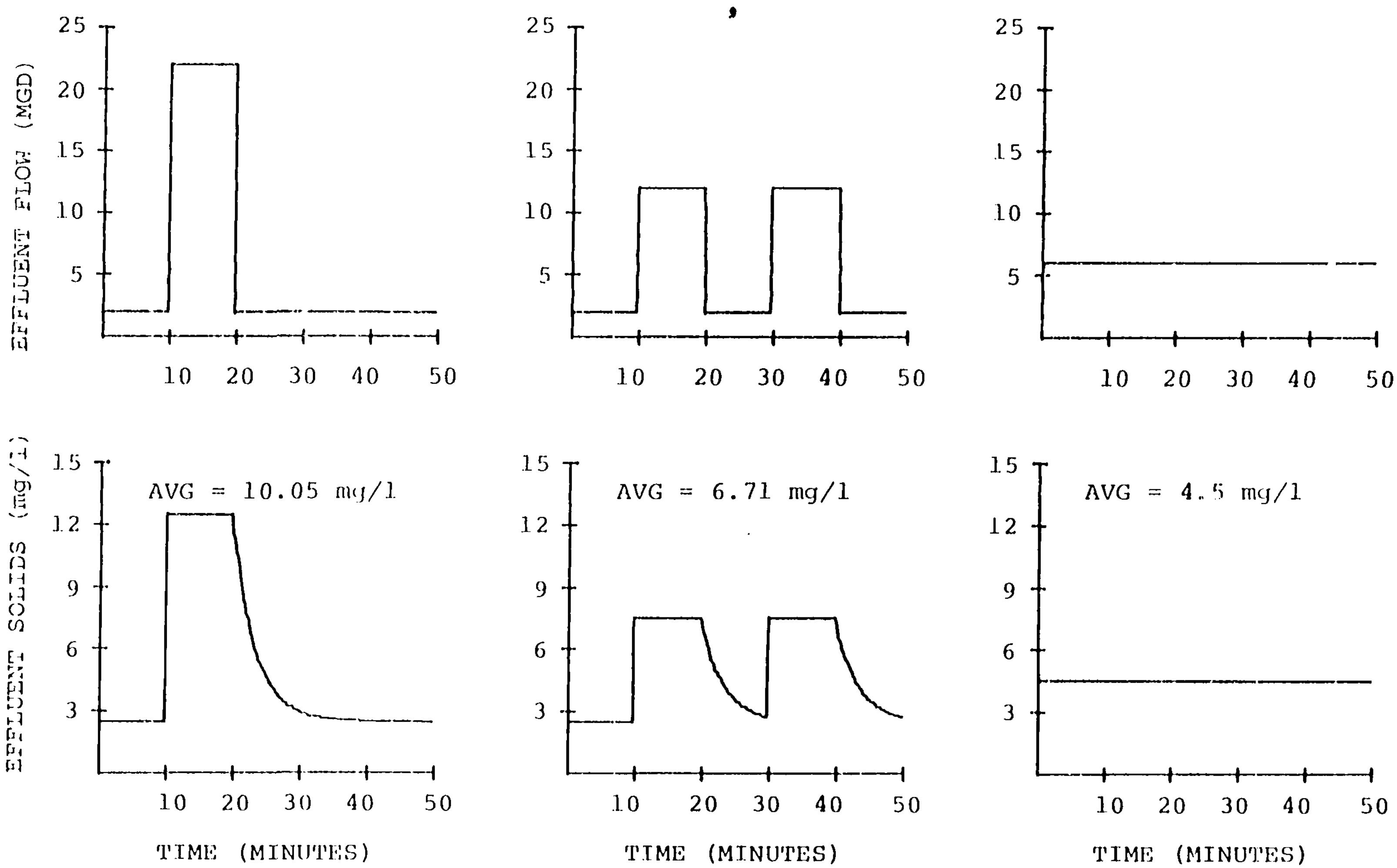

FIGURE 5.1 COMPARISON OF EFFECT OF THREE INFLUENT FLOW PATTERNS ON EFFLUENT SOLIDS

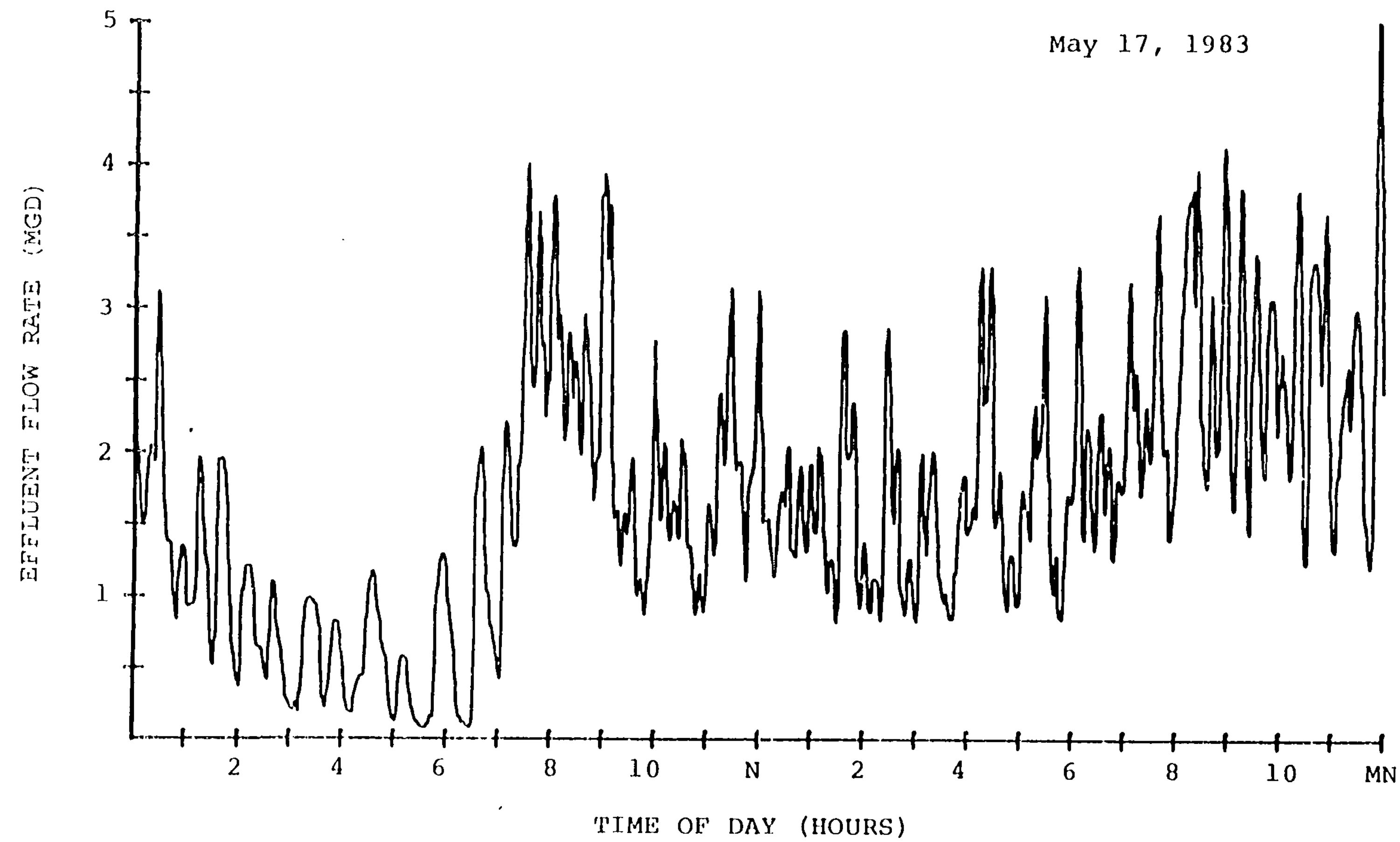

FIGURE 5.2 IRREGULAR EFFLUENT FLOW RATE PATTERN

aeration, and the reluctance of state and federal agencies to fund such processes. Many treatment plants (including Sagemont), however, do have some tankage volume which can be used for flow smoothing and equalization. This is the wet well and gravity collection system connected with the pumps station(s). What is required is a system of pumps for the variable flow pumping and an algorithm to control the pumps.

The Sagemont plant has a variable frequency pump and a pump with a modulating discharge valve for obtaining variable flow pumping. From a process point of view, either control method is acceptable.

At the Sagemont plant the wet well alone has only about 38,000 gallons (144,000 l) of useful storage or about 22 minutes of detention time at 2.5 MGD (9.4 M l/day). However, it is known from experience that it is possible to turn the influent pumps completely off for 4 to 6 hours before the collection system starts overflowing. This means that there is approximately another 500,000 gallons (1.9 M l) of storage in the gravity collection system draining to the plant.

Many operating engineers have been reluctant to use the storage capacity in the collection system. The primary objection to this usage relates to the deterioration of the sewer pipe itself. If sewage is held too long under anaerobic conditions, hydrogen sulfide is produced. The hydrogen sulfide is utilized by sulfur bacteria which produce sulfuric acid as a product. The sulfuric acid, in turn, is very corrosive to sewer pipe (especially concrete pipe) and tends to slowly destroy the crown of the pipe. Any operating scheme which holds the sewage in the collection system longer than necessary risks accelerating this process. The actual risk, however, is difficult to estimate since it depends on the retention time in the collection system, DO concentrations in the sewage, and temperature of the wastewater. Fortunately this risk can be minimized by pumping the wet well down several times a day. Other methods (131, 132) of controlling sewage sulfides are also available.

Even with only a limited wet well volume, it is possible to eliminate much of the rapid flow fluctuations by changing the pump station control strategy. One such control strategy is to control the wet well level at a constant value by varying the flow rate. The flow pattern resulting from such a control scheme is shown in Figure 5.3. This control strategy should indeed improve the discharge of effluent solids. However, there are still a number of sharp peaks in the flow curve due to other lift stations pumping into the on-site pump station. Some of these variations could be eliminated by utilizing a more intelligent pump controller.

Formulation of Flow Optimization Problem.

The proposed control strategy for influent pumping is based upon flow control with constraints on the wet well level and utilizes a dynamic optimization technique. The use of a process control computer is required. This technique is illustrated in Figure 5.4. For the arbitrary input shown in 5.4 (a), the optimum output is a constant flow as shown in 5.4 (b). However, with a finite equalization volume less than that needed for a constant flow, there will exist an optimal

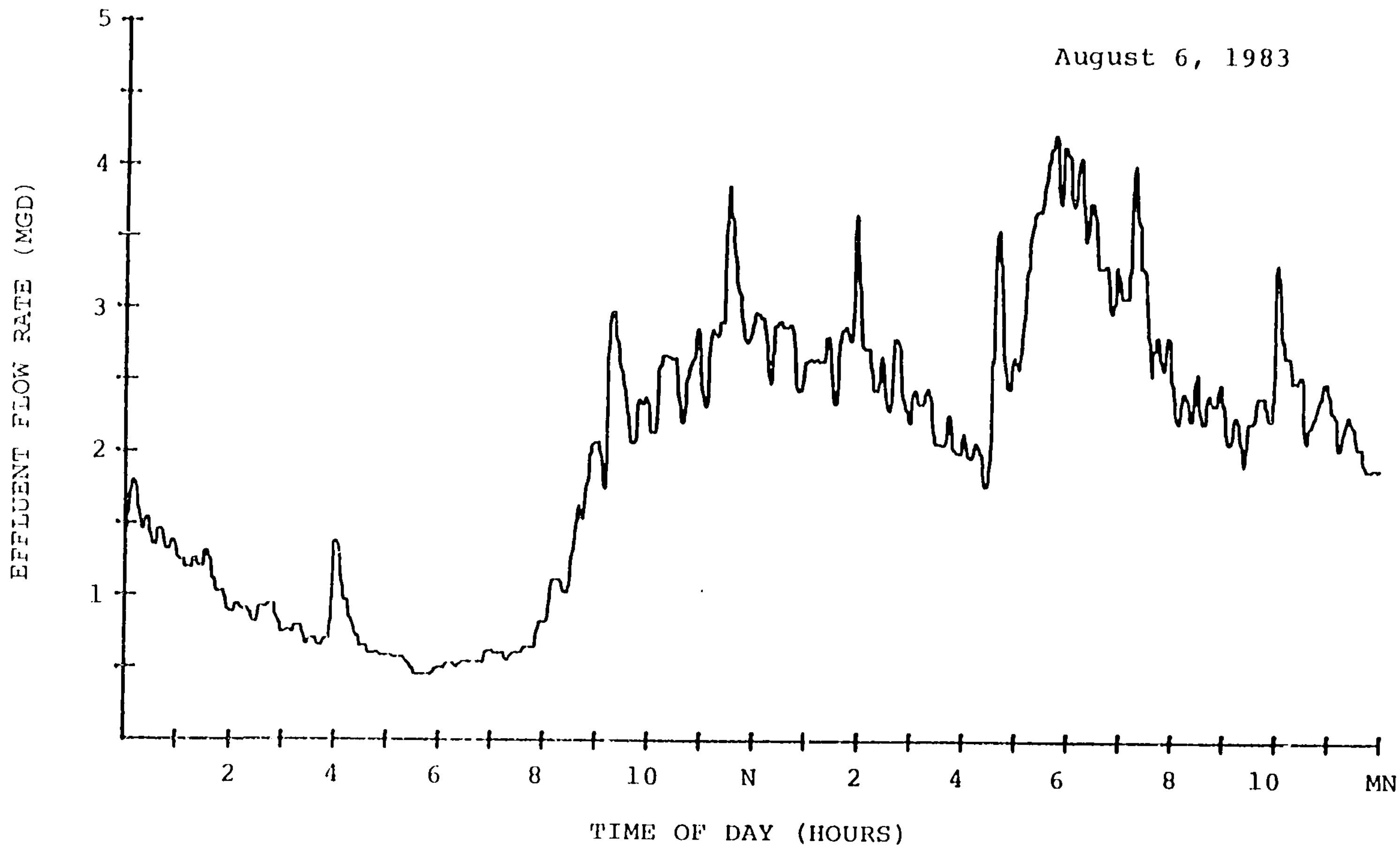

FIGURE 5.3 EFFLUENT FLOW RATE PATTERN WITH CONSTANT LEVEL CONTROL

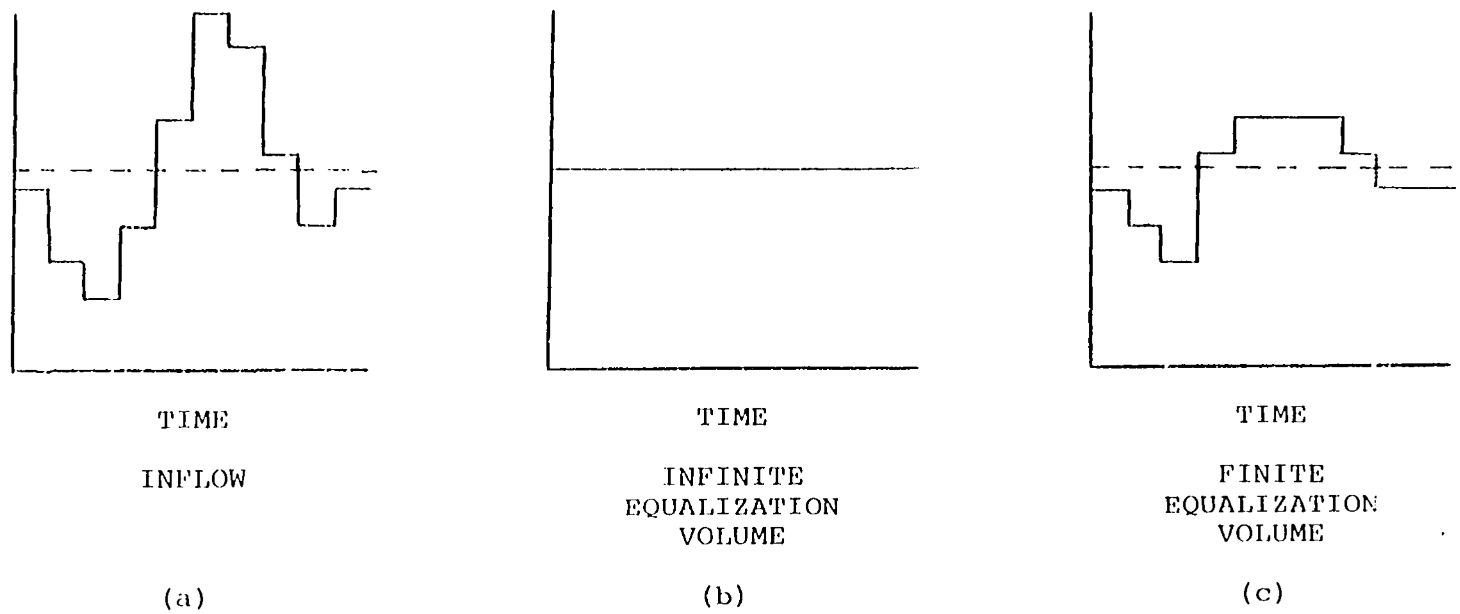

FIGURE 5.4 EFFECT OF EQUALIZATION VOLUME ON OPTIMAL FLOW PATTERN

outflow which minimizes the effluent solids discharge without overflowing or draining the equalization basin. This optimization problem is stated mathematically as Equation [5.2] with the constraints of Equations [5.3] and [5.4].

$$\text{minimize:} \quad \int_0^t F \cdot Xe \, dt \qquad [5.2]$$

$$\text{constraints:} \quad 0 < F < Fmax \qquad [5.3]$$

$$Hmin < H < Hmax \qquad [5.4]$$

where:

t = planning horizon (T),

F = outflow (L^3/T),

Xe = suspended solids concentration (M/L^3),

$Fmax$ = Maximum outflow (L^3/T),

$Hmin$ = minimum acceptable wet well level (L),

H = wet well level (L).

$Hmax$ = maximum acceptable wet well level (L).

The solution to this optimization problem itself is not of much use for control purposes since it would be too late to exert any control after all the flows were measured. Thus, for use as a control strategy, the flows into the plant must be predicted. Due to the stochastic nature of the inflow, however, these predictions will usually have some error. Therefore it is always necessary to update the optimization as new information is received. The flow forecasting technique developed in this investigation and dynamic optimization technique used are discussed in the following sections.

Flow Forecasting

Figure 5.5 shows five consecutive days of flow data (15 minute averages). All the days show a similar pattern although no two days are identical. A number of alternatives are available for forecasting of wastewater flow rates. Box and Jenkins (12) have developed a time-series technique for such analysis. This technique was applied by Goel and LaGrega (56) specifically for forecasting wastewater flows. Beck (8) also used a stochastic model derived from time-series analysis with adaptive prediction to forecast flows. Olsson (86), however, has suggested a simpler method. The regularity of the flows from day to day suggests that the flow could be forecast based upon an historical average and some type of correction term based upon recent flow values as in Equation [5.5].

$$\hat{f} = \bar{f} + fc \qquad [5.5]$$

where:

$\hat{f}$ = flow prediction (L^3/T),

$\bar{f}$ = historical average flow (L^3/T),

fc = correction term (L^3/T).

The average of the five days of flow from Figure 5.5 are shown in Figure 5.6. This function can be approximated in several manners including interpolation between the data points, Fourier transform analysis, and time-series analysis methods. A simpler alternative to the time-series technique is Fourier transform analysis (140). It is particularly appropriate for describing periodic functions although it is generally considered to be a less powerful tool than time-series

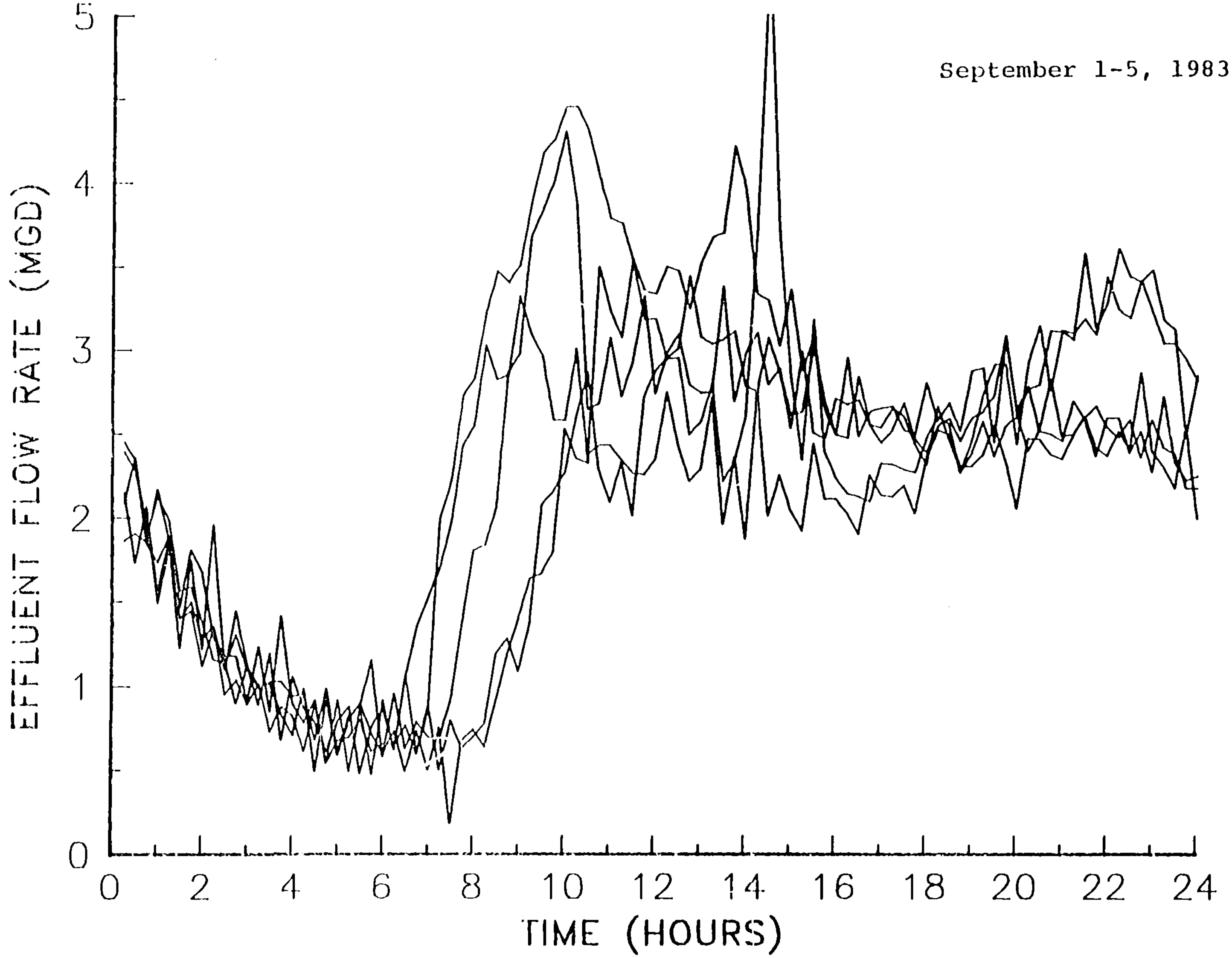

FIGURE 5.5 FIVE CONSECUTIVE DAYS OF INFLUENT FLOW DATA

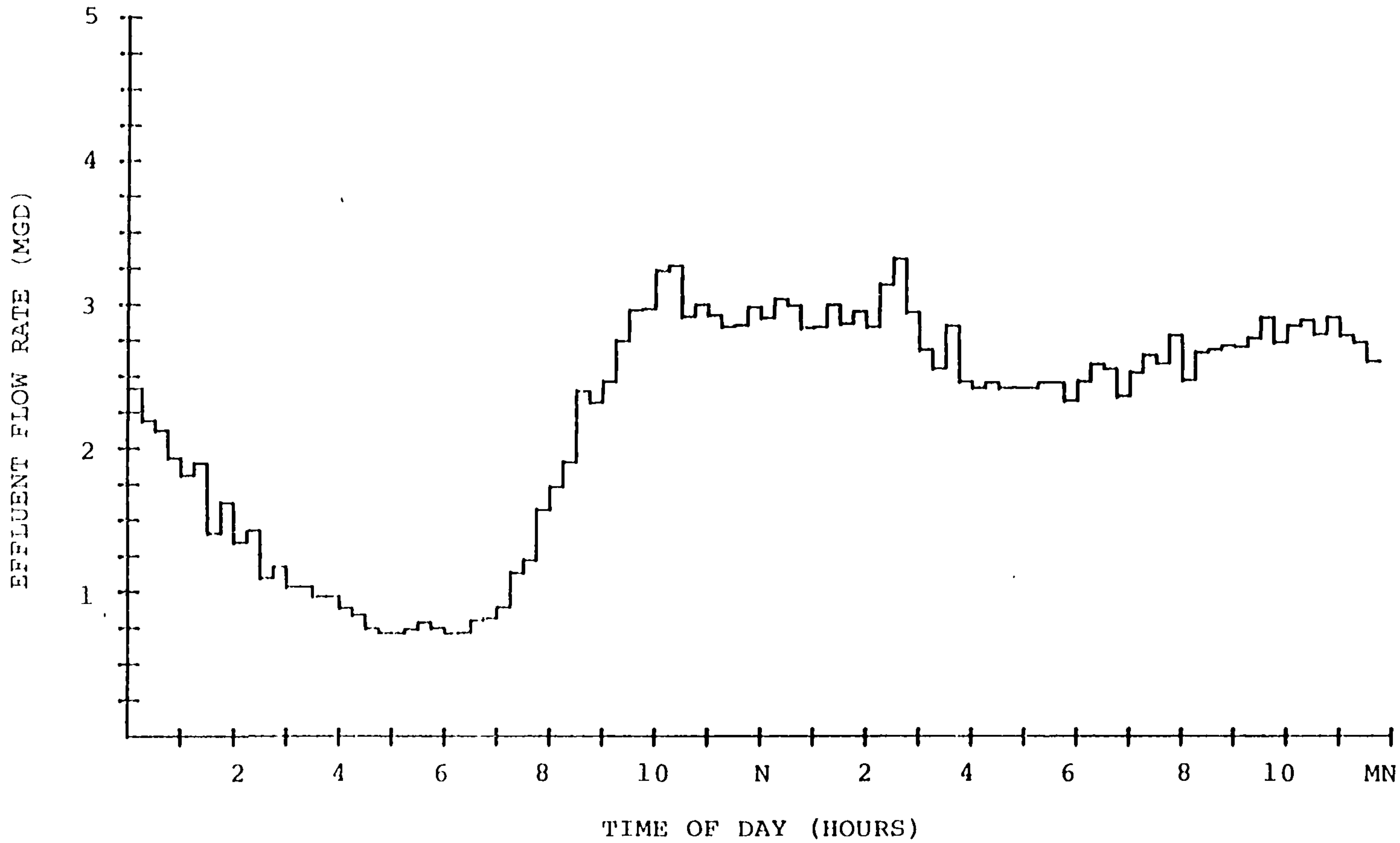

FIGURE 5.6 HISTORICAL AVERAGE FLOW USED IN INFLUENT FLOW PREDICTOR

analysis (110). Equation [5.6] shows the form of the Fourier series approximation.

$$\bar{f}(t) = \sum_{i=0}^{n} a(i)\cos(2\cdot i\cdot PI\cdot t/T) + b(i)\sin(2\cdot i\cdot PI\cdot t/T) \qquad [5.6]$$

where:

$a(i), b(i)$ = Fourier series coefficients,

PI = 3.14159

t = time (T),

T = period (T).

The Fourier coefficients may be calculated by a number of methods. Most of the Fast Fourier Transform (FFT) techniques require a fixed sampling frequency. The program used in this investigation was adapted from a microcomputer utility manual (65) and rewritten in FORTRAN. It allows the use of unequally spaced data. It was found that six coefficients resulted in a good approximation of the average flow data as shown in Figure 5.7. Numerical values for these coefficients are listed in Table 5.1.

The correction term, fc, can be approximated with a polynomial of the following form:

$$fc(t) = \sum_{i=1}^{m} c(i)\cdot[F(t-i) - \bar{f}(t-i)] \qquad [5.7]$$

where:

$c(i)$ = polynomial coefficients,

$F(t)$ = actual measured flow (L^3/T).

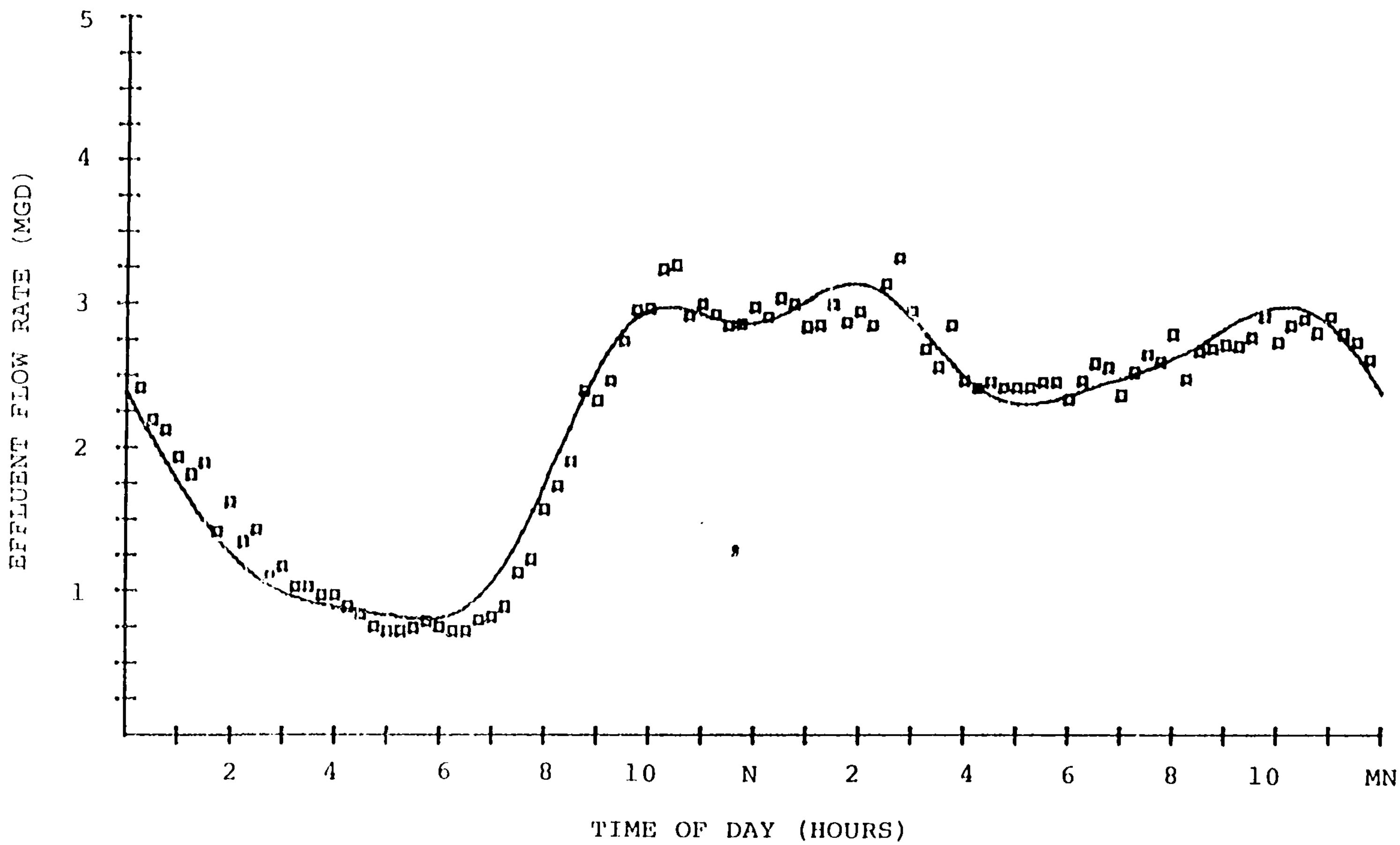

FIGURE 5.7 FOURIER APPROXIMATION OF FLOW DATA

TABLE 5.1

FOURIER COEFFICIENTS FOR FLOW APPROXIMATION

i	a(i)	b(i)
0	2.202161	0.0
1	-0.4042748	-0.7756295
2	0.5608708	-0.3562915
3	0.03840571	0.0568838
4	-0.1031323	-0.0809117
5	0.1246335	-0.05210978
6	-0.04647651	0.006037218

It has been found that four coefficients in Equation [5.7] result in a satisfactory approximation. The coefficients for a 15 minute predictor are listed in Table 5.2. Different coefficients can be derived for 30, 45, 60, etc. minute predictors. Alternately, Equation [5.5] can be applied consecutively with the 15 minute coefficients using the predicted flows in place of the missing measurements. Each method results in equally good predictions. Obviously, predictions further in the future show more error than the shorter term ones.

Dynamic Optimization

The control of the influent flow is determined by making an optimal decision at each of a sequence of points in time. The interval between successive decision points is called a period or stage (102). It is assumed that a decision is made at the beginning of each period. The time interval over which the optimization is made consists of N periods and is called the planning horizon. Both the length of the period and the planning horizon are determined by system constraints.

The length of time between decision points is based on several criteria. To maximize the effects of control, it is desirable to make the period as small as possible and "fine tune" the system. On the other hand, to minimize calculations it is desirable to make the period as large as possible. A practical matter is that many motors are rated for 3 to 4 starts per hour. This limits the period to 15 to 20 minutes or longer. Additionally, the wet well has a detention time of 22 minutes at average flow. It is desirable to have a period less than this. A period of 15 minutes was chosen as a compromise of these considerations.

TABLE 5.2

CORRECTION POLYNOMIAL COEFFICIENTS

i	c(i)
1	0.826906
2	0.194541
3	0.053819
4	-0.096410

The planning horizon is also determined as a compromise of several criteria and depends on the intended purpose of the control. For instance, if a large equalization volume is available and the intent is to equalize daily flow variations, the planning horizon should be large, perhaps 24 hours. However, if only a small equalization volume is available and the intent is to minimize short-term flow fluctuations, a shorter planning horizon is appropriate. For the purpose of this investigation, planning horizons on the order of one to two hours are considered.

Error Function

The optimization technique requires a single response for each set of inputs. Typically this response is some type of error function which is evaluated from the given inputs. Since the Simplex method used in this investigation attempts to maximize the response and the desired result is to minimize the error, the response is taken as the negative of the total error.

$$R = -E_T \qquad [5.8]$$

where:

R = response,

E_T = total error.

The Simplex method is an unconstrained optimization technique. To satisfy the constraints (Equations [5.3] and [5.4]) of the problem, it is necessary to include a penalty error term which has a large value when these constraints are violated. Thus the total error is composed of a term reflecting the discharge of effluent solids and a penalty term reflecting the state of the constraints.

$$E_T = Ex + Ep \qquad [5.9]$$

where:

Ex = error term due to discharge of solids,

Ep = penalty error term due to constraints.

Others (41) have included further error terms to eliminate rapid changes in outflow. This inclusion is considered unnecessary in the present work since rapid changes in flow will be reflected in the solids discharge term.

Ideally the solids discharge error term, Ex, would reflect the true integral of discharged solids as in Equation [5.2]. This would require that the full dynamic hydraulic model coupled with the decay model for clarification (Equations [4.14], [4.15], and [4.16]) be solved for 1 to 2 hours of simulated time for each response. Since many responses are needed for the decision at each period, this would lead to excessive computational requirements. It is quite possible that a small process control computer would not be capable of doing the required computations in real-time.

An alternative to the full-blown dynamic model can be derived by making a few simplifying assumptions. Since the change in flow will be relatively slow (compared to on/off operation of the influent pumps) and the plant has relatively little hydraulic dampening capacity, it is reasonable to assume the effluent flow rate is approximately equal to the influent flow rate. This assumption eliminates the need to use the dynamic hydraulic model for the optimization response. Additionally, if the flow is not changing rapidly, it is reasonable to use the simplified clarification model (Equation [4.14] only) rather than the full decay model. These simplifications reduce each evaluation to solving one differential equation (for wet well level) and a number of algebraic equations.

Thus the error term due to the discharge of solids, Ex, has the form:

$$Ex = \sum_{i=1}^{n} F(t+i) \cdot Xe(t+i) \qquad [5.10]$$

where:

n = number of period in planning horizon,

$Xe(t) = K_1 + K_2 \cdot F(t)$,

K_1, K_2 = model parameters from Equation [4.14].

The penalty error term, Ep, is zero when no constraints are violated and takes a large value for each constraint violation. Each constraint (Equations [5.3] and [5.4]) is treated separately.

$$Ep = Ep(1) + Ep(2) + Ep(3) \qquad [5.11]$$

where:

Ep(1) = penalty error term due to flow constraint violations,

Ep(2) = penalty error term due to low wet well levels,

Ep(3) = penalty error term due to high wet well levels.

The error term due to flow constraint violations, Ep(1), is considered a hard constraint since it is not physically possible to exceed the limitations of the pumps. It takes the form:

$$Ep(1) = \sum_{i=1}^{n} 1000 \cdot [U(-F(t+i)) + U(Fmax-F(t+i))] \qquad [5.12]$$

where:

U(t) = step function = 0 if t<0 and 1 if t>0

The low wet well level limit is also considered a hard constraint since equipment could be damaged if the pumps were allowed to run dry. However, this penalty term also includes a linearly increasing part to differentiate between the severity of violations.

$$Ep(2) = \sum_{i=1}^{n} U(Hmin-H(t+i)) \cdot [1000+100 \cdot (Hmin-H(t+i)] \qquad [5.13]$$

where:

H(t) = wet well level at time t (L).

Economically, it might be desired to keep the wet well level as high as possible to reduce energy costs for pumping. This feature could be incorporated by including another error term for low wet well level. This feature was not incorporated.

The high wet well limit is considered a soft limit since it is permissible for the sewage to back up into the collection system during some periods of the day. Additionally, a very soft constraint was established to maintain a small buffer for unexpected short-term transients of the inflow.

$$Ep(3) = \sum_{i=1}^{n} U(H(t+i)-Hsoft)\cdot[1.0\cdot(H(t+i)-Hsoft)] +$$

$$U(H(t+i)-Hmax)\cdot[100+100\cdot(H(t+i)-Hmax)] \qquad [5.14]$$

where:

Hsoft = soft limit level for wet well (L).

VI. EVALUATION OF PROCESS CONTROL STRATEGIES

Influent Flow Rate Control

In the previous chapter a dynamic optimization technique was developed for minimizing the effluent suspended solids concentration by shaping the influent flow utilizing the buffering capacity of the wet well volume. This section evaluates this control strategy with traditional on/off control and constant level control strategies by computer simulation. The simulation language CSMP (66) was originally used for these simulations. The model was rewritten into the ACSL (80) language for the final simulations.

The sizes and capacities of the lift station, pumps, reactors, and settler were taken from the Sagemont plant. The settler model parameters, however, were estimated from data presented by Chapman (21). Chapman's data were used since they were considered more typical of most wastewater treatment plants than the parameters associated with a plant operating as well as the Sagemont WWTF. In order to test the control strategy over a wide range of conditions, one day of normal, dry weather flow and one day of extreme storm flow were chosen for evaluation. These data are shown in Figures 6.1 and 6.2. One minute average flows were used for input to the simulations.

On/Off Controller

A simplified on/off controller was chosen for evaluation. The on/off controller model response is shown as Figure 6.3. When a pump is on,

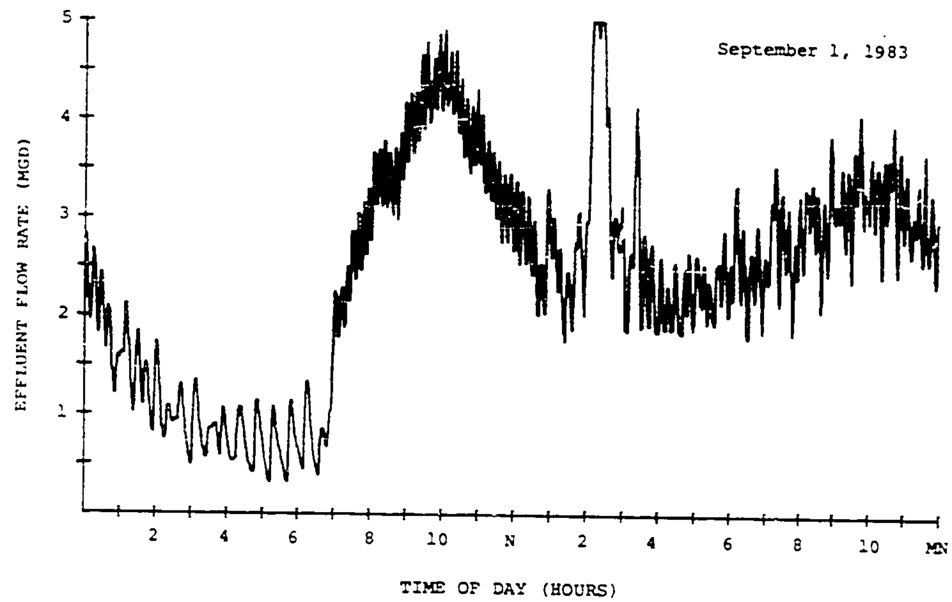

FIGURE 6.1 DRY WEATHER FLOW PATTERN

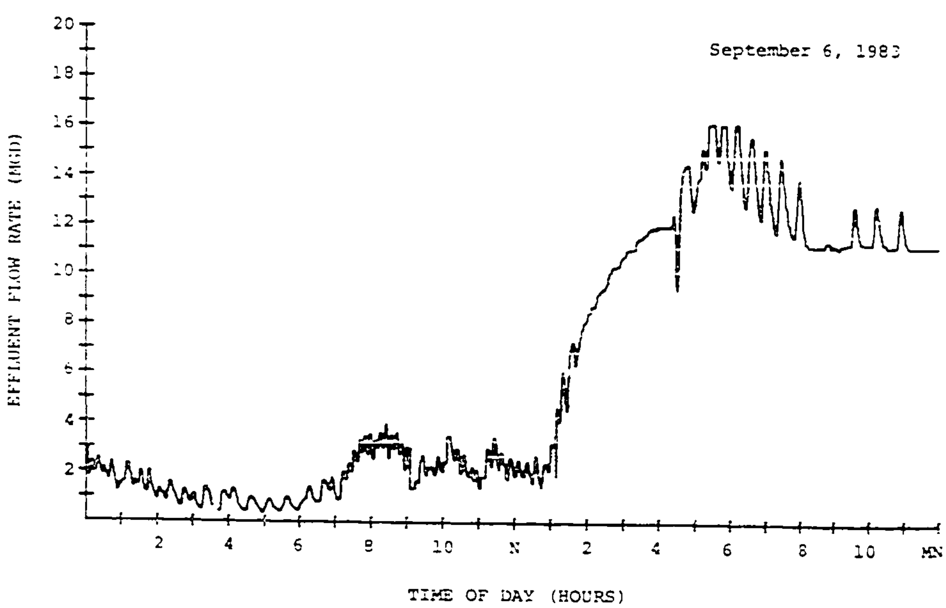

FIGURE 6.2 STORM FLOW PATTERN

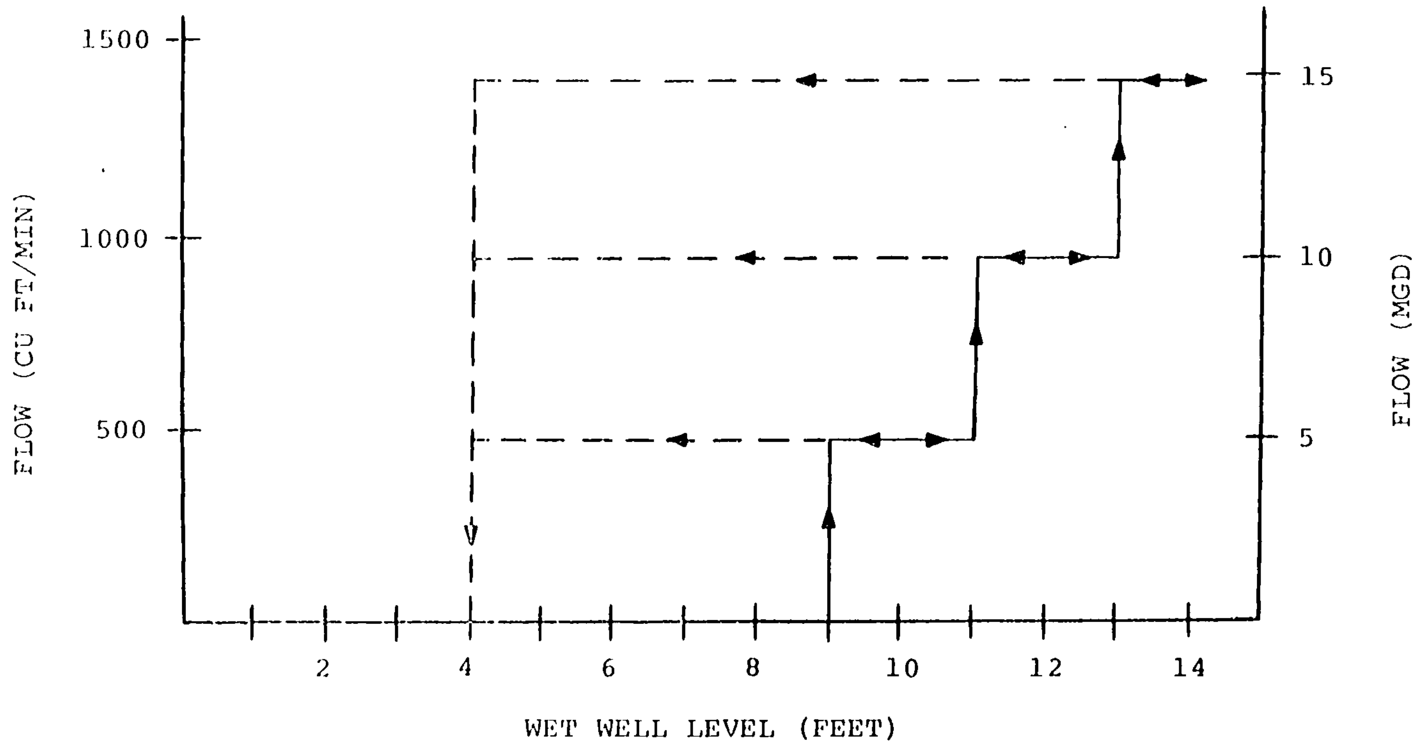

FIGURE 6.3 ON/OFF CONTROLLER ALGORITHM

the flow from that pump is assumed constant. This ignores the fact that the flow decreases as the wet well level drops. This assumption, however, introduces less than 5 percent error over the entire range studied.

For the dry weather flow, only one pump is ever running at any given time. For the storm flow, all three pumps run at some times.

Level Controller

The level control algorithm used in the evaluation is shown as Figure 6.4. This algorithm is similar to that implemented at the Sagemont plant. This algorithm represents an improvement over the on/off controller by controlling the flow proportional to the wet well level over a portion of its range.

Again, for the dry weather flow, only one pump is required. For the storm flow all three pumps are required.

Optimization Controller

The optimization controller developed in the previous section was used with a period of 15 minutes and planning horizons of one and two hours. Simulations with a two hour planning horizon showed no significant improvement over those using one hour. Therefore, the results for the one hour planning horizon will be presented in the remainder of this chapter. This is not a totally unexpected result since the detention time of the wet well was only 22 minutes. Had the volume for flow

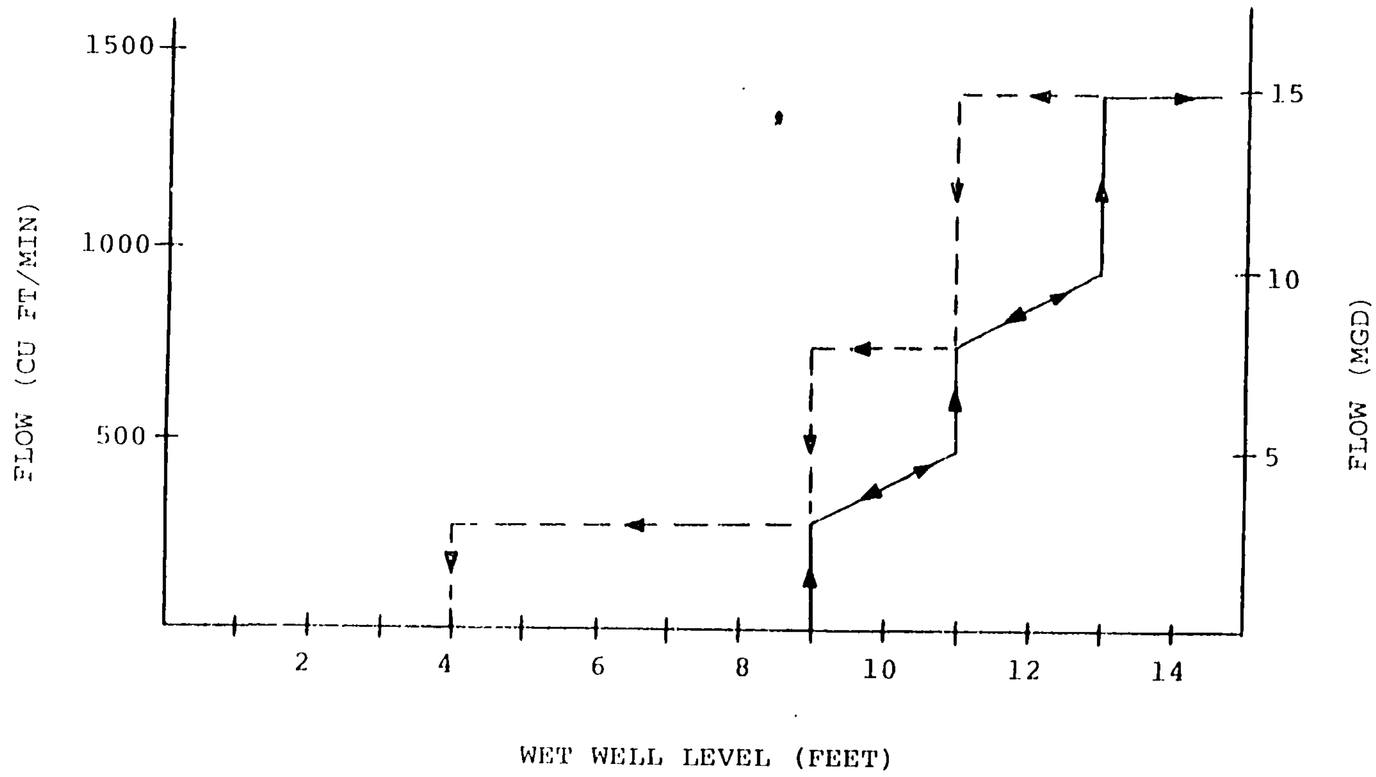

FIGURE 6.4 CONSTANT LEVEL PUMP CONTROL ALGORITHM

equalization been greater, the longer planning horizon should have shown an improvement.

Evaluation of Simulation Results

Table 6.1 shows the simulation results for the three control strategies for normal, dry weather flow. Table 6.2 shows the results for the storm flow. For the dry weather flow, both the level and optimization controllers showed significant improvements (11.7 and 12.1 percent respectively) over the on/off controller. For the storm flow, the level and optimization controllers showed more modest improvements of 3.6 and 3.2 percent respectively. For reference, simulations of the optimization controller with a "perfect" flow predictor were also conducted. These simulations show that further improvements (up to 15.9 and 5.4 percent respectively) are possible with refined flow prediction.

Although improvements of 15.9 percent might be considered small, it should be noted that these are specific for the assumed parameters and plant studied. Improvements larger than those noted in this study might be possible for overloaded plants.

Two factors contribute to the differences in improvements between dry weather and storm flow conditions. With the larger flow, the detention time (and hence buffering capacity) of the wet well is reduced leading to less chance for improvement by control. Also, for the storm flow the average flow was 87 percent of the capacity of one pump compared to 34 percent for the dry weather flow. Thus, just by the magnitude of the

TABLE 6.1

SIMULATION RESULTS OF
INFLUENT FLOW CONTROL STRATEGY
FOR DRY WEATHER FLOWS

CONTROL METHOD	24-HOUR AVERAGE S.S. CONCENTRATION	PERCENT IMPROVEMENT OVER ON-OFF CONTROL
ON/OFF CONTROL	19.6	---
LEVEL CONTROL	17.3	11.7
OPTIMIZATION CONTROL	17.2	12.1
OPTIMIZATION CONTROL[1]	16.5	15.9

[1] - with "perfect" flow predictor

TABLE 6.2

SIMULATION RESULTS OF
INFLUENT FLOW CONTROL STRATEGY
FOR STORM FLOWS

CONTROL METHOD	24-HOUR AVERAGE S.S. CONCENTRATION	PERCENT IMPROVEMENT OVER ON/OFF CONTROL
ON/OFF CONTROL	49.9	---
LEVEL CONTROL	48.2	3.6
OPTIMIZATION CONTROL	48.4	3.2
OPTIMIZATON CONTROL[1]	47.3	5.4

[1] - with "perfect" flow predictor

flows, the capacity of the on/off pump was closer to that for optimal control.

In summary it can be said that the level controller gives improvement in effluent suspended solids when compared to the on/off controller. Under most conditions, the optimization controller makes further improvements by utilizing the wet well volume to equalize and stabilize the flow. The amount of improvement depends both upon the magnitude and distribution of the inflow.

It should be noted that the optimization controller only adjusted the outflow once per 15 minutes. The simulations with the on/off controller and the level controller used a one minute period. Additional simulations have shown the on/off and level controllers to be unstable for periods in the range of 5 to 10 minutes and longer.

Feed Forward Flow Control

In the previous section it was proposed that feed forward flow could be used to rapidly reduce the MLSS concentration entering the settler. As shown in Figures 4.9 (June 25, 1983) and 4.10 (July 27, 1983), two experiments were conducted to demonstrate this phenomenon. In each case, the MLSS concentration decreased in the manner expected. This change is shown in greater detail in Figure 6.5. In this test, the MLSS concentration decreased from 6,000 to 4,000 mg/l in approximately 75 minutes from the start of the forcing. This represents a 33 percent reduction in solids loading to the settler.

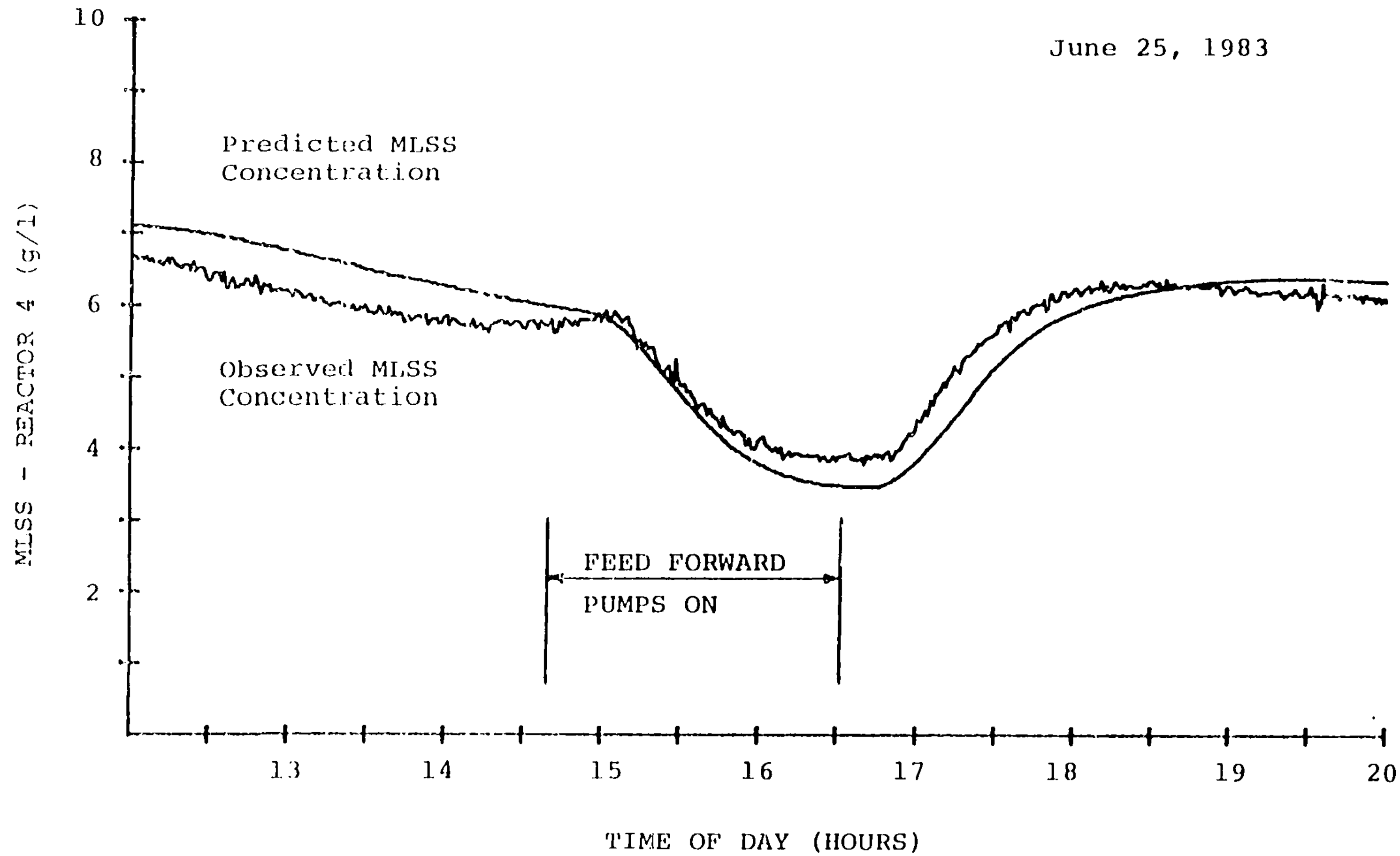

FIGURE 6.5 MLSS RESPONSE TO FEED FORWARD FLOW - BASIN 4

Figure 6.5 also shows the simulation results for the last reactor using the reactor mixing model for the same change in feed forward flow. The model shows good correlation with experimental data with a maximum error of about 8 percent. Figure 6.6 shows the simulation results and experimental data for the same forcing in reactor 2. Even though the variation is less, the simulation again shows good correlation.

Return Sludge Flow Control

In the previous section, proportional control of the recycle flow rate was proposed to minimize variations of the MLSS and return sludge concentrations. Experimental data from the Sagemont plant show these phenomena.

Data from April 15, 1983 are shown in Figures 6.7, 6.8, and 6.9. For this period, the return sludge flow rate was held constant at a value of approximately 150 percent of the average flow. For this type of control, the MLSS concentration varied from 5,200 to 6,200 mg/l. The return sludge concentration varied from 7,000 to 12,000 mg/l and was approximately in phase with the influent flow. Problems associated with these variations are discussed in a later section.

Data from April 12, 1983 are shown in Figures 6.10, 6.11, and 6.12. For this period, the return sludge flow rate was varied proportional to a exponentially filtered value of the effluent flow rate. The set point was 100 percent. For this type of control, the MLSS concentration remained nearly constant. The return sludge concentration also remained more nearly constant over a large portion of the day. The exception to

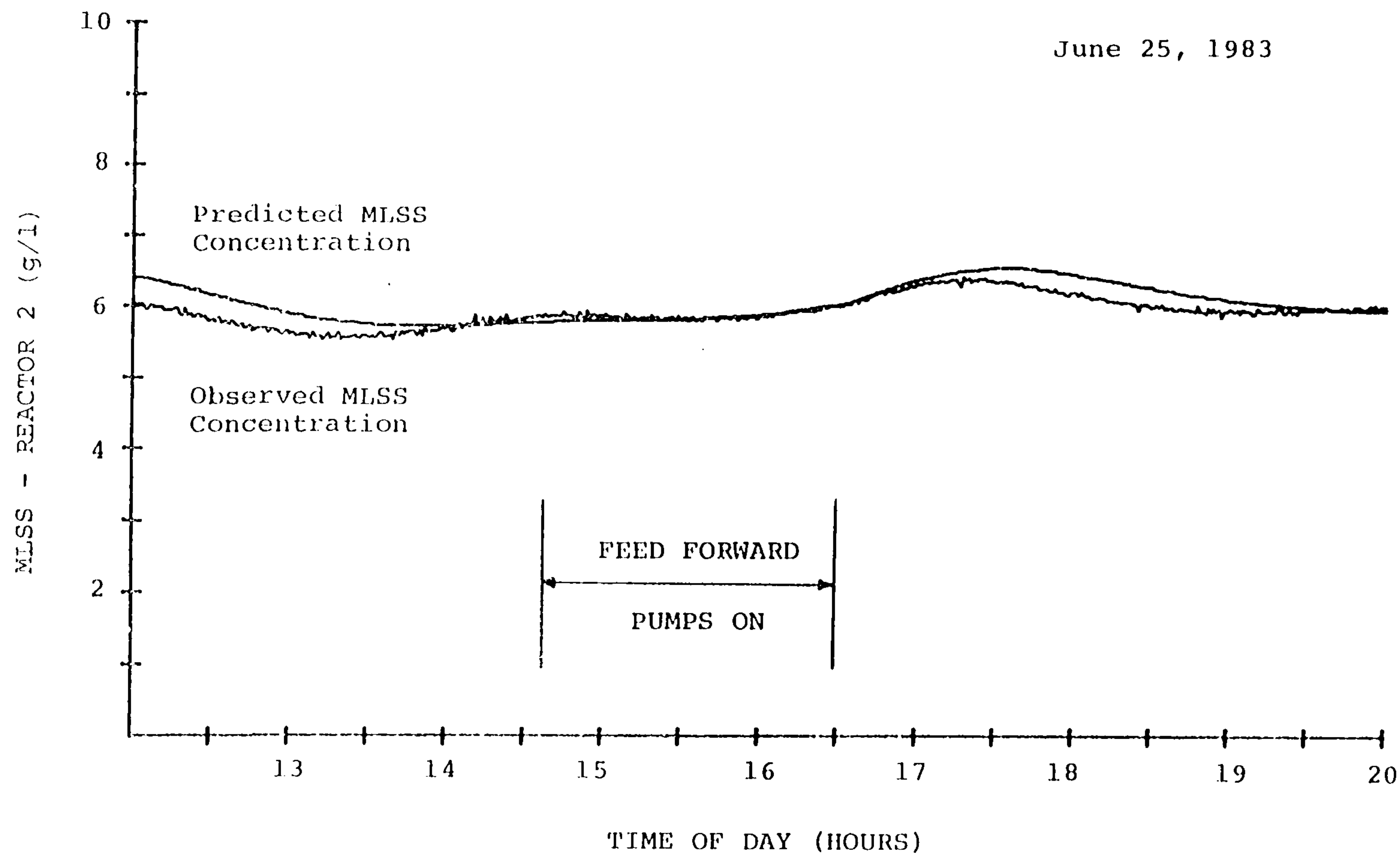

FIGURE 6.6 MLSS RESPONSE TO FEED FORWARD FLOW - BASIN 2

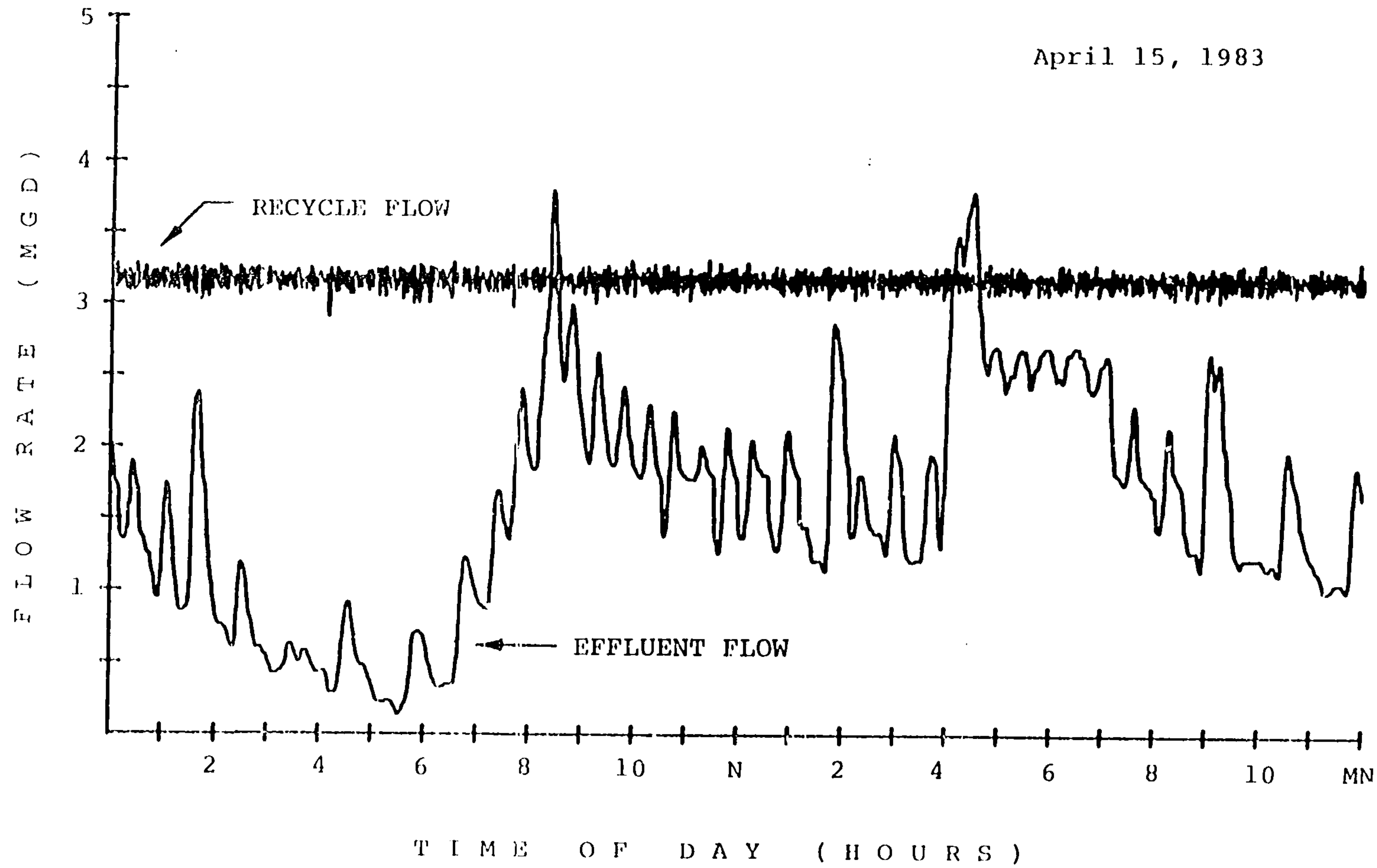

FIGURE 6.7 EFFLUENT FLOW AND CONSTANT RECYCLE FLOW RATE CONTROL

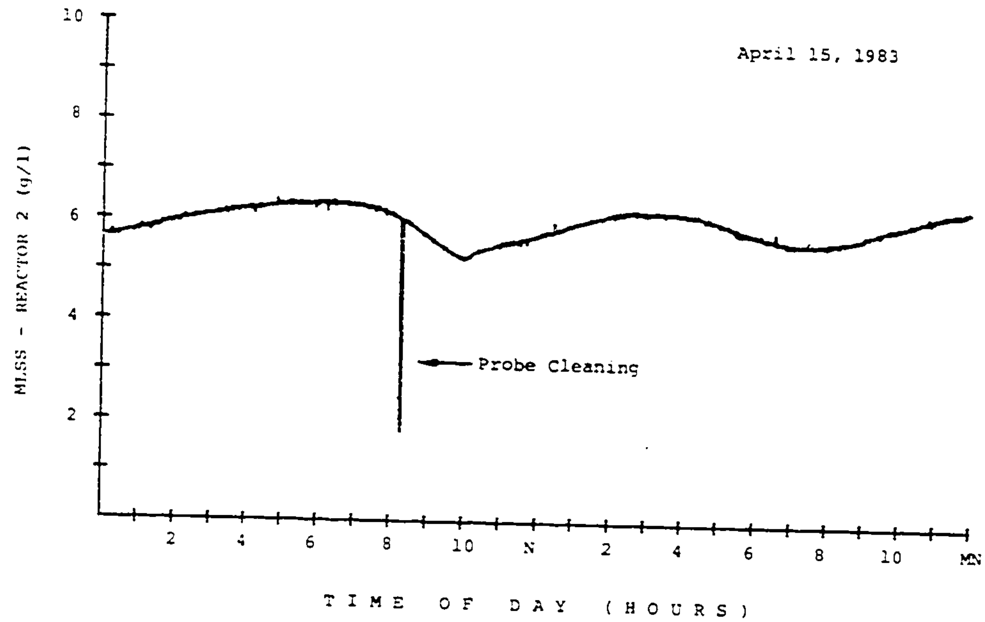

FIGURE 6.8 MLSS CONCENTRATION WITH CONSTANT RECYCLE FLOW RATE CONTROL

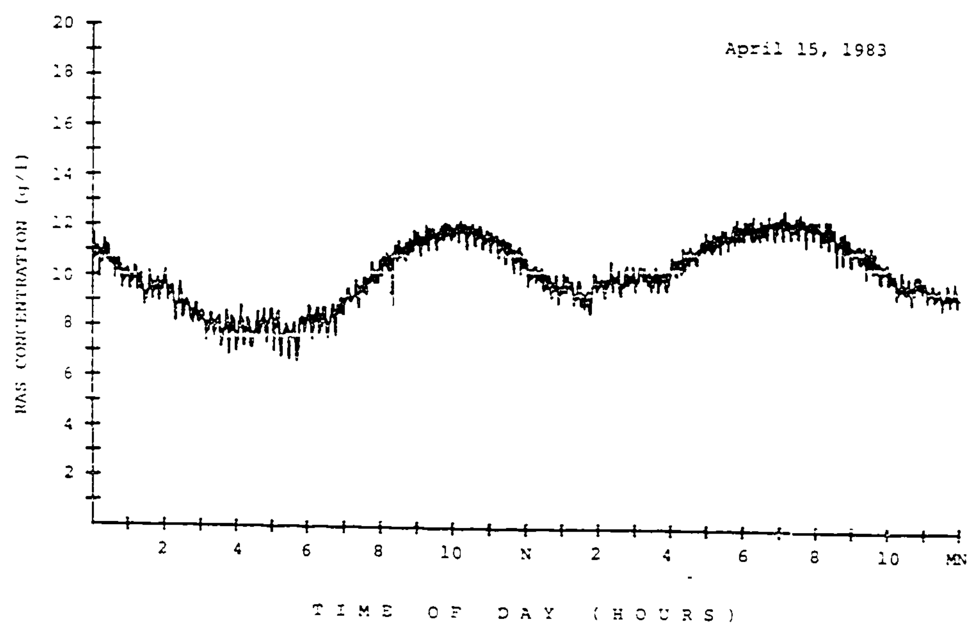

FIGURE 6.9 RETURN ACTIVATED SLUDGE (RAS) CONCENTRATION WITH CONSTANT RECYCLE FLOW RATE CONTROL

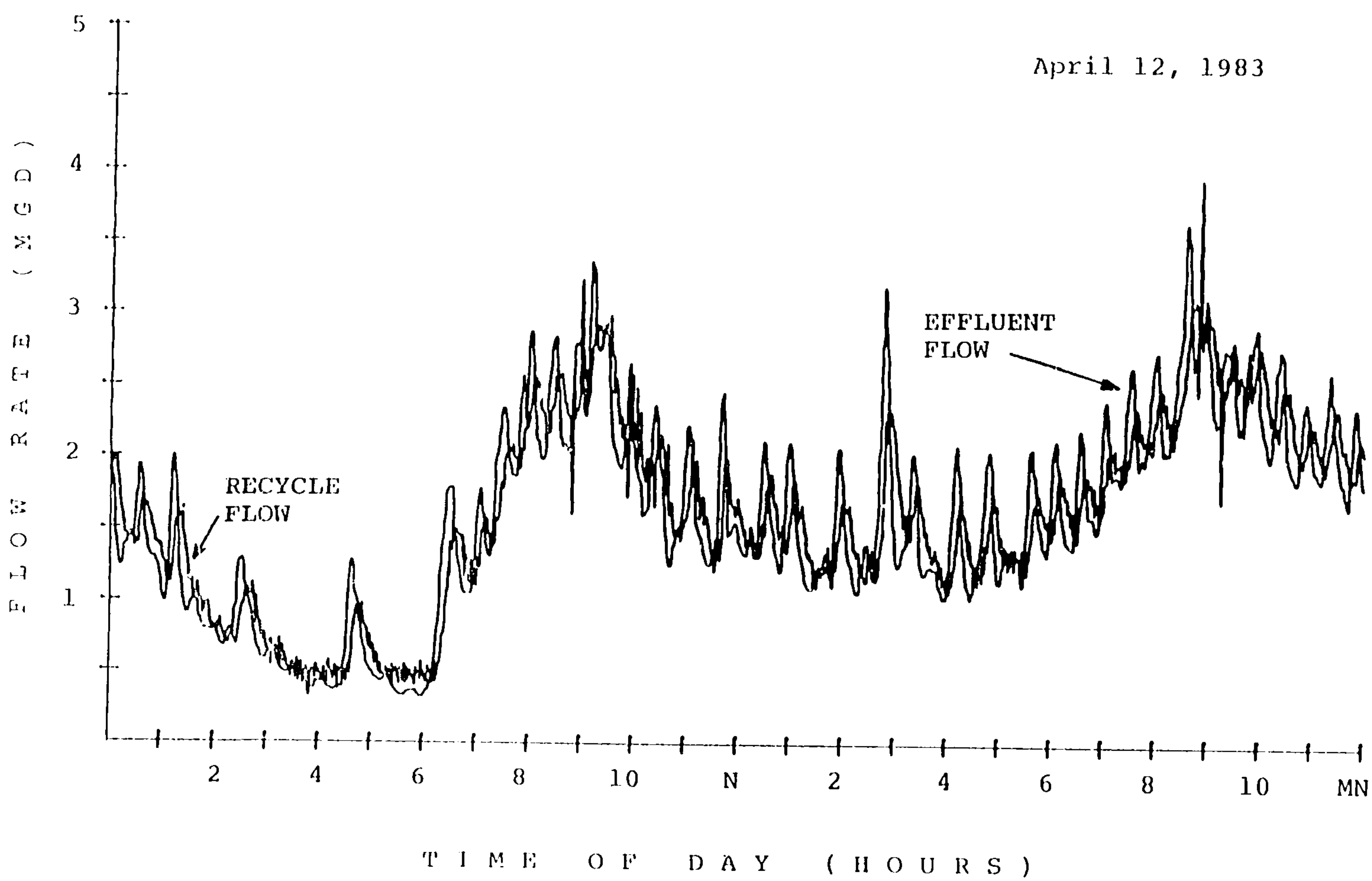

FIGURE 6.10 EFFLUENT FLOW AND PROPORTIONAL RECYCLE FLOW RATE CONTROL

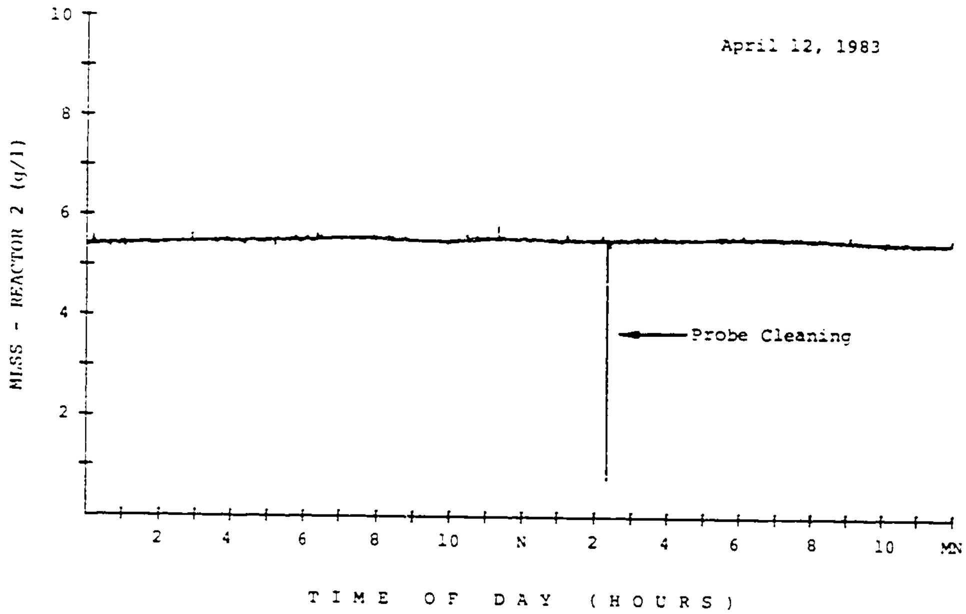

FIGURE 6.11 MLSS CONCENTRATION WITH PROPORTIONAL RECYCLE FLOW RATE CONTROL

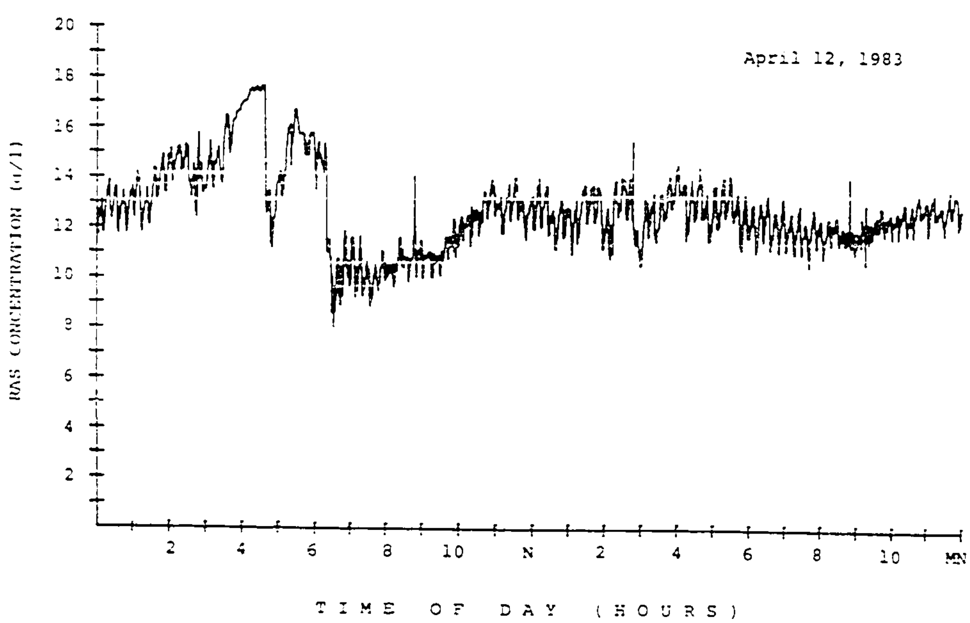

FIGURE 6.12 RETURN ACTIVATED SLUDGE (RAS) CONCENERATION WITH PROPORTIONAL RECYCLE FLOW RATE CONTROL

this occurred in the early morning hours and was a result of the low flows at that time and the wasting of excess sludge.

VII. ON-LINE STATE/PARAMETER ESTIMATION

Biological wastewater treatment processes are generally not as easily modeled as many of the processes from the chemical industries. This is due to a large number of factors including a heterogeneous biomass, shifting biological populations, and changing influent characteristics. With these conditions, models for biological processes are necessarily imperfect and variations in the parameters for these models should be expected. For these models, techniques for determining the variations of the model parameters and states is especially important.

As demonstrated in the previous sections, the parameters for the clarification and thickening models are not true constants; they vary with time. Therefore, in order to fully utilize the models for real-time control, some method must be used to update or adjust the model parameters so that the model reflects the true process. Depending on the process, several alternatives are available for this estimation. If natural forcings are sufficient and the parameters vary slowly, it may be sufficient to store process measurements for a period (say one day) and perform an off-line analysis to calculate new parameters. If natural forcings are not sufficient for a good estimation, it may be necessary to purposely perturb the process and estimate parameters from that data. For processes where the parameters change more rapidly, it may be advantageous to estimate them with a recursive estimation technique.

Recursive Estimation

The recursive estimation technique utilized in this investigation was presented by Olsson (88). The technique is capable of estimating both process states and parameters and can be considered equivalent to a simplified Kalman filter. A brief review is included below to demonstrate the technique.

Consider the case of an inert dye continuously injected into two CSTRs in series. The influent concentration is assumed to be known, the effluent concentration is measured, and it is desired to estimate the dye concentration in the first CSTR. The mass balances describing the system are:

$$d(c_1)/dt = -(F/V)\cdot c_1 + (F/V)\cdot c_0 \qquad [7.1]$$

$$d(c_2)/dt = -(F/V)\cdot c_1 - (F/V)\cdot c_2 \qquad [7.2]$$

where:

c_0, c_1, c_2 = dye concentrations (M/L^3),

F = flow rate (L^3/T),

V = reactor volume (L^3).

Equations [7.1] and [7.2] can be written compactly in matrix form as shown in Equation [7.3].

$$dX/dt = A \cdot X + B \cdot u \qquad [7.3]$$

where:

$$X = \begin{bmatrix} c_1 \\ c_2 \end{bmatrix} \qquad A = \begin{bmatrix} -F/V & 0 \\ F/V & -F/V \end{bmatrix} \qquad B = \begin{bmatrix} F/V \\ 0 \end{bmatrix} \qquad u = c_0$$

Given accurate influent measurements and a perfect model, Equation [7.3] will eventually converge to the correct concentration values. This method, however, is slow and ignores the process measurement, c_2. It should be possible to use the measurements to speed up the rate of convergence. Olsson postulated that a way to utilize process measurements to adjust the model is to add a correction term to Equation [7.3].

$$d\hat{X}/dt = A \bullet \hat{X} + B \bullet u + K \bullet (y - C \bullet \hat{X}) \qquad [7.4]$$

where:

$\hat{X}$ = model states,

K = correction matrix, constant,

y = process measurements, $C \bullet X$,

C = relationship of measured value to states, (0 1).

Equation [7.4] can be generalized as:

$$d\hat{X}/dt = f(\hat{X},u) + K \bullet (y - C \bullet \hat{X}) \qquad [7.5]$$

For linear systems, it can be shown that the estimation error using Equation [7.5] will converge to zero providing that the model and measurements are perfect. For nonlinear models this convergence cannot be guaranteed. However, when initial guesses are reasonable, the estimator generally behaves well.

Up to this point only the estimation of states has been considered. The technique can be expanded to estimate model parameters as well. Assume that it is desired to estimate a model parameter $\hat{a}_i$. a_i is introduced as a new process variable using the differential equation:

$$d(a_1)/dt = 0.0 \qquad [7.6]$$

where:

a_1 = process parameter.

In the estimator, the equation for $\hat{a}_1$ would be:

$$d(\hat{a}_1)/dt = 0.0 + K\cdot(y-C\cdot\hat{X}) \qquad [7.7]$$

where:

$\hat{a}_1$ = model parameter.

Besides being able to estimate time-varying model parameters, this technique can also be useful in simplifying the models used for control. Often complex models can be approximated with less complex models and time-varying parameters with adequate results.

Olsson gives only general guidelines in the selection of the estimator adjustment matrix, K. Usually the sign of the matrix coefficients can be determined a priori by a careful analysis of the model. However, their values depend on the absolute values of the states and parameters being estimated, the confidence in the model, and the noise level of the measurements. If there is great confidence in the model and the measurements are noisy, the matrix coefficients should have relatively small values such that the model states and parameters are adjusted slowly. Likewise, if the measurements are very accurate, the coefficients should have larger values to fully account for the additional information the process measurements give.

Application to the Clarification Model

As noted, one major application of recursive estimation is to reduce the complexity of the models needed for control. In this application, the model defined by Equation [4.10] is used rather than the full decay model (Equations [4.14], [4.15], and [4.16]). The equation is repeated below for convenience.

$$Xe = K_1 + K_2 \cdot Fe \tag{4.10}$$

In the estimator, it is assumed that all variation is contained in the parameter K_2. This is reasonable since Table 4.4 showed that K_1 was relatively constant in many of the experiments. Additionally, it was not felt necessary to adjust the state (effluent solids), only the parameter. Thus the model equations become:

$$\hat{Xe} = \hat{K}_1 + \hat{K}_2 \cdot Fe \tag{7.8}$$

$$d(\hat{K}_2)/dt = 0.0 + K_{est} \cdot (Xe-\hat{X}) \tag{7.9}$$

where:

K_{est} = estimator coefficient.

The new parameter K_{est} can be assigned almost any (positive) value. If made large enough, the parameter K_2 will vary fast, and the model prediction will almost exactly match the measured values. A more rational method of determining its value is demonstrated in Figure 7.1. In this figure, the instantaneous values of K_2 are plotted as a function of time and a trend line has been fit. The time varying values of K_2 obtained with various values of K_{est} are also shown. From examination

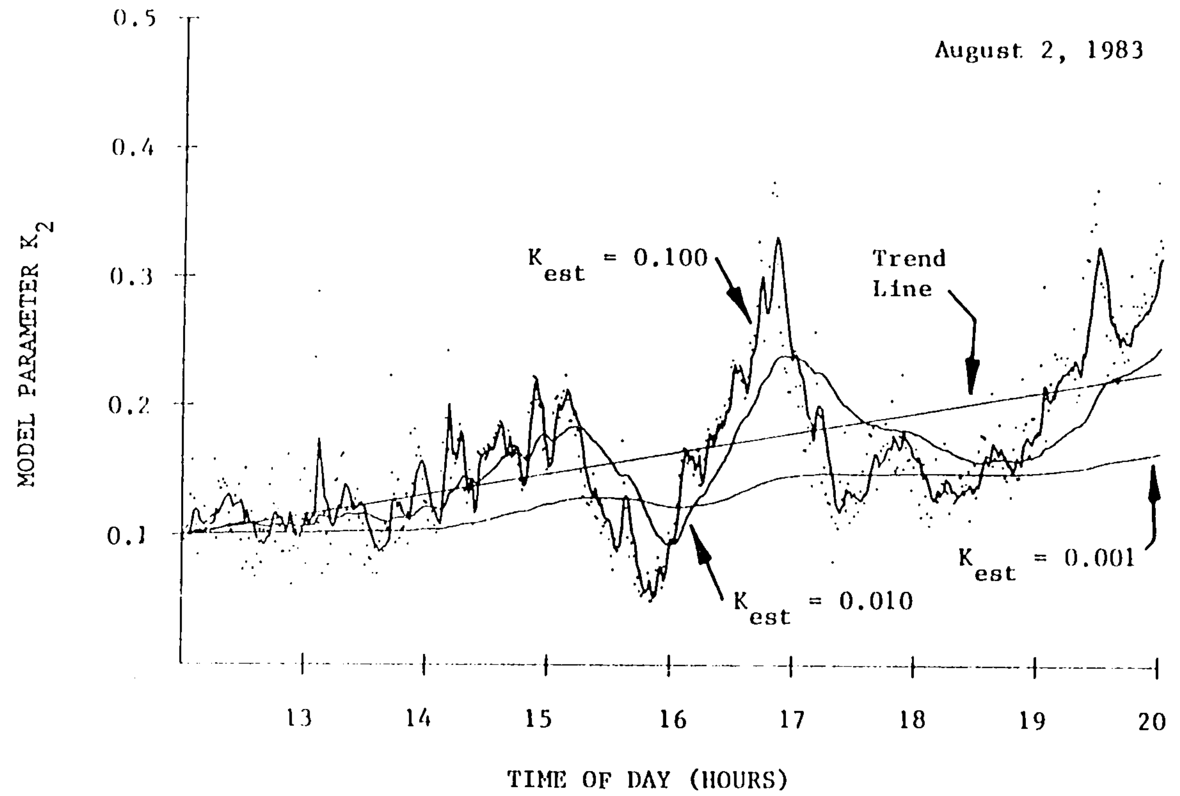

FIGURE 7.1 EFFECT OF ESTIMATOR COEFFICIENT ON RATE OF CONVERGENCE FOR MODEL PARAMETER K_2

of several such plots, it was determined that a value for K_{est} of 0.005 to 0.010 was a good compromise using one minute flow and effluent solids data. Values greater than this allowed too much variation of K_2 while values less than this did not allow K_2 to vary fast enough to account for the observed variations.

Results of simulations based upon the recursive model are shown in Table 7.1 and plotted in Figures 7.2, 7.3, and 7.4. It should be noted that the results of the recursive model compare well to those of the decay model proposed in section IV.

<u>Application to the Thickening Model</u>

The recursive estimation technique can also be applied to the settler thickener model developed in section IV to estimate both states (solid concentrations in each element) and the parameters in the settling velocity function (Eq. [4.21]). To apply the state estimation, each of the thickening model equations must be modified to include a correction term. These equations are:

$$d\hat{X}_1/dt = (FLUXIN - Fr \cdot \hat{X}_1)/V_1 - A_1 \cdot MIN(\hat{G}s_1, \hat{G}s_2)/V_1 + K_{est} \cdot (X_1 - \hat{X}_1) \qquad [7.10]$$

$$d\hat{X}_i/dt = Fr \cdot (\hat{X}_{i-1} - \hat{X}_i)/V_i + A_{i-1} \cdot MIN(\hat{G}s_{i-1}, \hat{G}s_i)/V_i - A_i \cdot MIN(\hat{G}s_i, \hat{G}s_{i+1})/V_i + K_{est} \cdot (X_i - \hat{X}_i) \qquad [7.11]$$

$$d\hat{X}_n/dt = Fr \cdot (\hat{X}_{n-1} - \hat{X}_n)/V_n + A_{n-1} \cdot \hat{G}s_n)/V_n + K_{est} \cdot (X_n - \hat{X}_n) \qquad [7.12]$$

TABLE 7.1

PARAMETERS FOR EQUATIONS [7.8] AND [7.9]

DATE	K_i	K_{EST}	NUMBER DATA POINTS	SUM RESIDUAL SQUARED	AVG ERROR
April 21, 1983	0.78	0.010	480	38.2	0.28
April 23, 1983	0.16	0.010	285	50.3	0.42
April 26, 1983	1.68	0.010	480	162.3	0.58
August 2, 1983	1.74	0.010	480	43.0	0.30
August 3, 1983	1.79	0.010	480	23.1	0.22
August 7, 1983	1.81	0.010	160	24.9	0.39

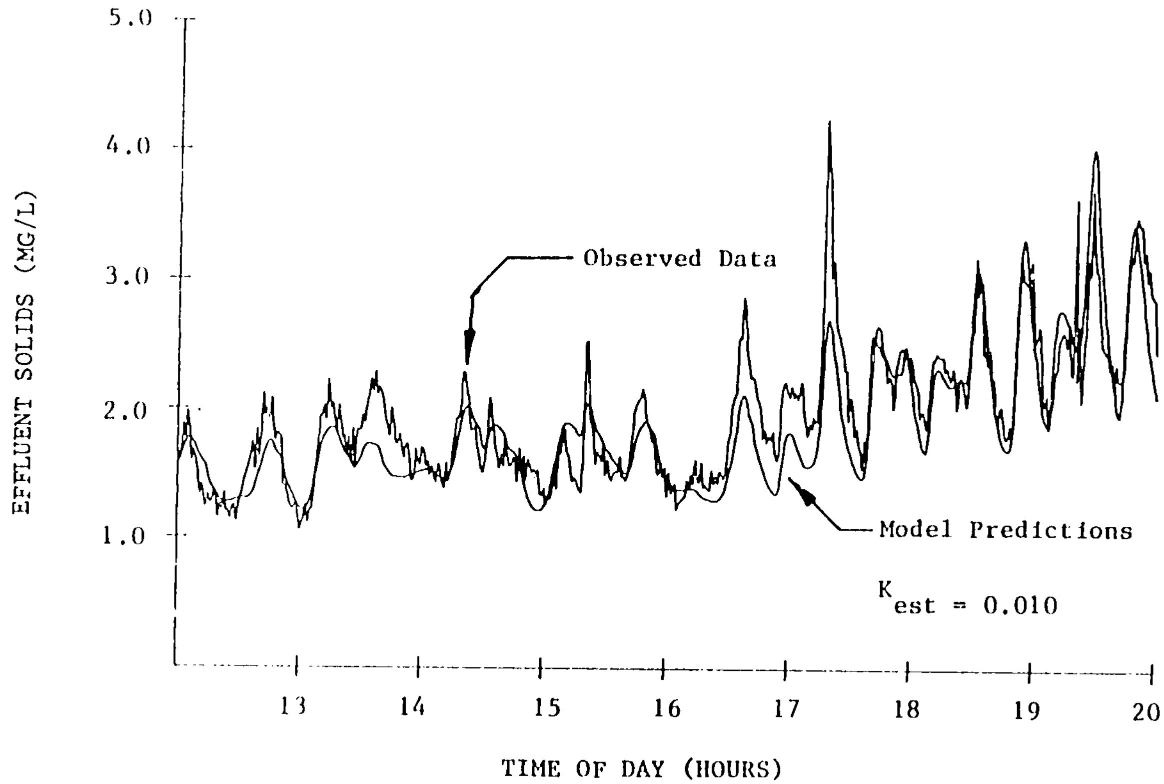

FIGURE 7.2 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATIONS [7.8] AND [7.9] - APRIL 21, 1983

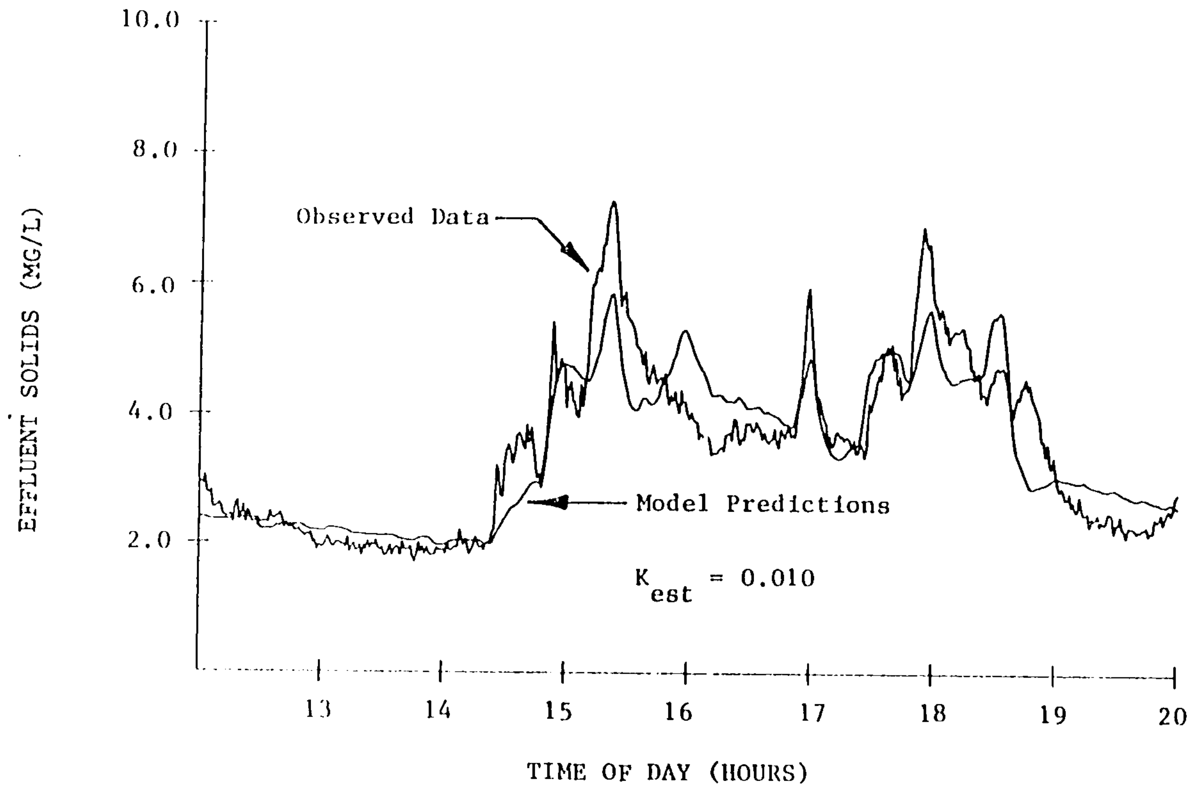

FIGURE 7.3 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATIONS [7.8] AND [7.9] - APRIL 26, 1983

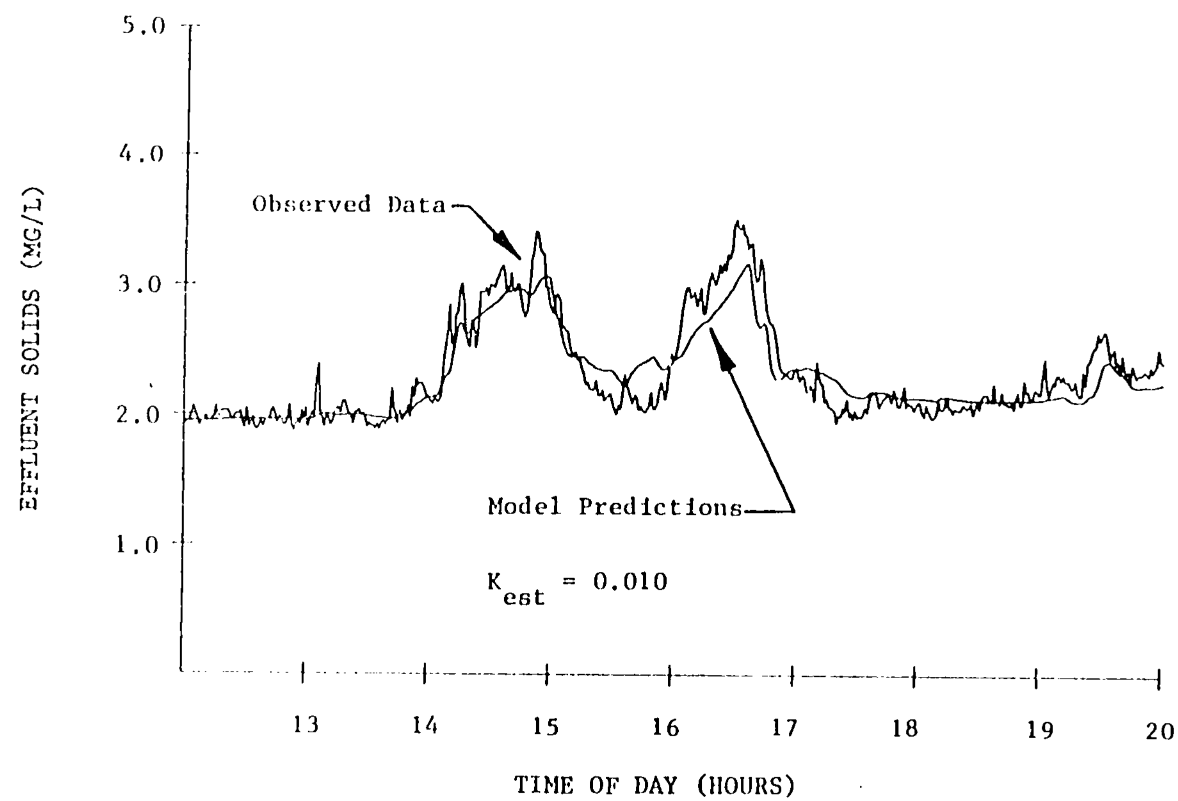

FIGURE 7.4 EFFLUENT SUSPENDED SOLIDS PREDICTIONS BASED UPON EQUATIONS [7.8] AND [7.9] - AUGUST 2, 1983

The absolute value for K_{est} can be estimated from simulations as shown in Figure 7.5. In this example, correct values for the settling parameters and incorrect initial conditions are assumed. Different values of K_{est} lead to varying rates of correction as the solids concentrations come to steady state values. Even with a value of K_{est} of zero (no correction term), the states will eventually converge to the correct values. However, this convergence is very slow. Generally, values of K_{est} between 1.0 and 10.0 lead to satisfactory results.

The recursive estimation of the two parameters in the settling velocity function (Eq. [4.21]) is somewhat more difficult than that of the states since the settling velocity must be inferred from changes in concentration rather than any direct measurement. This inference is further complicated by the fact that the change in concentration in any element in the thickener model is affected by the solids concentration in that element, the concentration in the element above it, and the concentration in the element below it. The two exceptions to this are the bottom element and the top element of the sludge blanket zone. The bottom element has no layer beneath it while the top element of the sludge blanket zone has only a low concentration layer above it, which has little effect.

Simulations have shown that the element where the sludge blanket begins (nearest the surface) is the most sensitive to changes in settling parameters. This is reasonable since all changes in these parameters must propagate through the entire sludge blanket before they show up at the bottom element. Therefore, the estimation equations for the

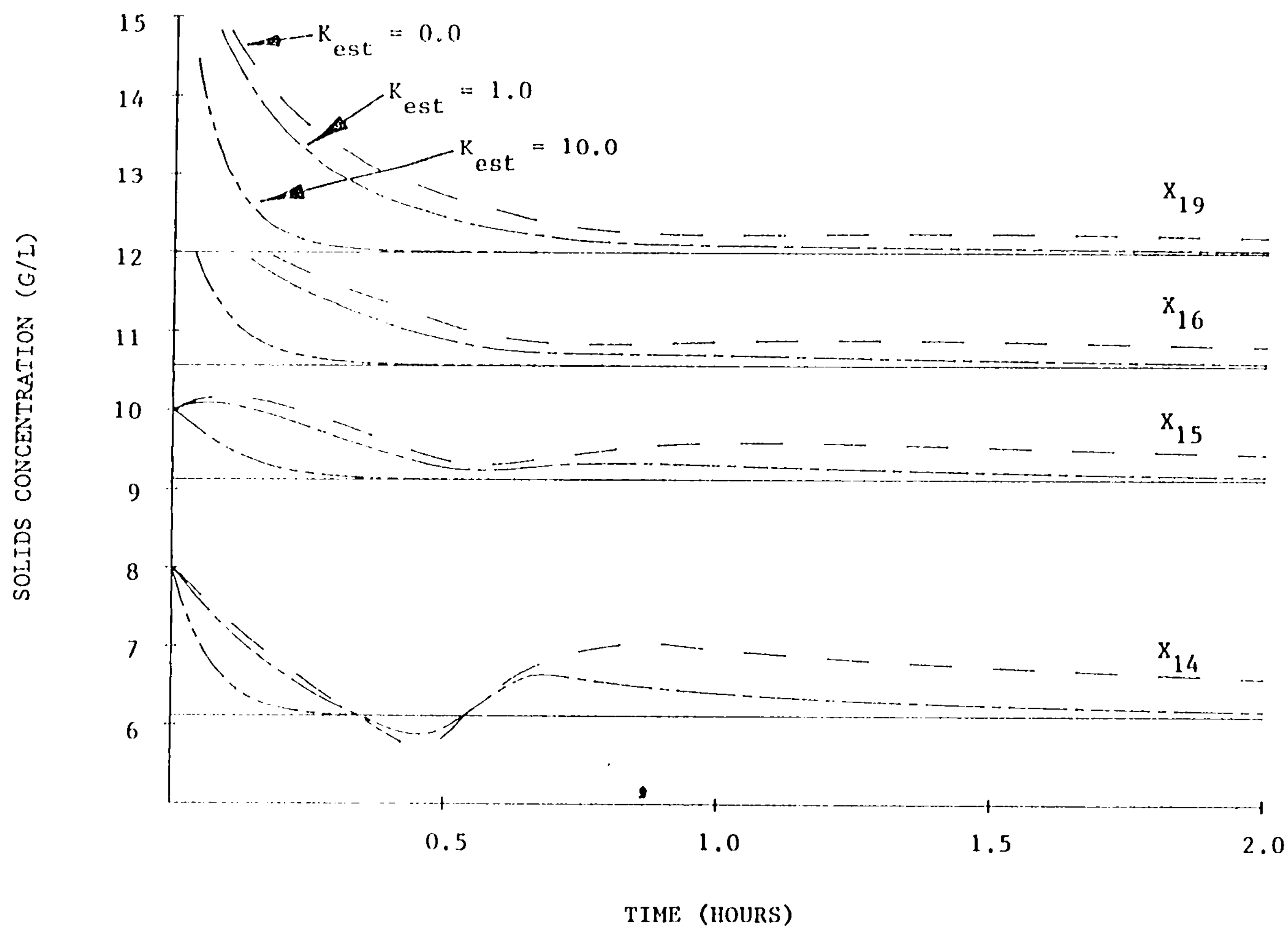

FIGURE 7.5 EFFECTS OF K_{est} ON RATE OF STATE ESTIMATION CONVERGENCE

settling parameters are based on the top blanket element and take the form:

$$d(\hat{V}_0)/dt = 0.0 + K_{V0} \cdot (X_{sb} - \hat{X}_{sb}) \qquad [7.13]$$

$$d(\hat{b})/dt = 0.0 + K_b \cdot (X_{sb} - \hat{X}_{sb}) \qquad [7.14]$$

where:

K_{V0} = estimator coefficient,

K_b = estimator coefficient,

X_{sb} = solids concentration in top element of sludge blanket zone, (M/L^3),

$\hat{X}_{sb}$ = estimated solids concentration in top element of sludge blanket zone (M/L^3).

Again, acceptable values of K_{V0} and K_b can be established through simulations. The simplest case involves estimating only one of the two parameters at a time. Figures 7.6 and 7.7 show the convergence of the the parameter $\hat{V}_o$ for different values of K_{V0} and K_{est} (state estimation) while Figures 7.8 and 7.9 show the convergence of the parameter $\hat{b}$ over a similar range. These simulations were conducted with the correct initial conditions and values of $\hat{V}_0$ and $\hat{b}$ which varied approximately ten percent from the "real" model. In general, the simulations with K_{est} of 10.0 (Figures 7.7 and 7.9) showed very rapid convergence of the settling parameters while the simulations with K_{est} of 1.0 (Figures 7.6 and 7.8) showed more oscillation and slower convergence. Acceptable values of K_{V0} are in the range of -0.02 to -0.05, and acceptable values of K_b are in the range of 3.0×10^{-7} to 1.0×10^{-6}.

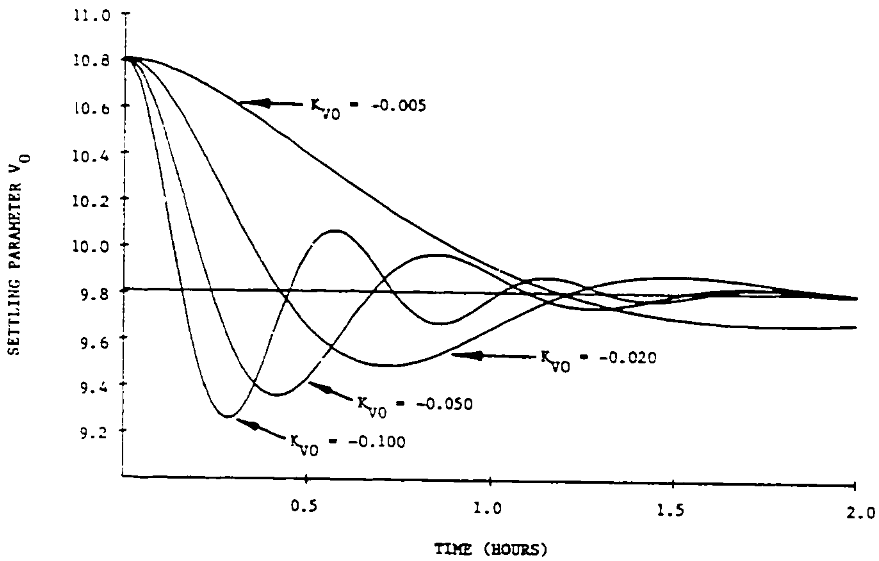

FIGURE 7.6 EFFECT OF ESTIMATOR COEFFICIENTS ON RATE OF CONVERGENCE OF SETTLING PARAMETER V_0 (K_{est}=1.0)

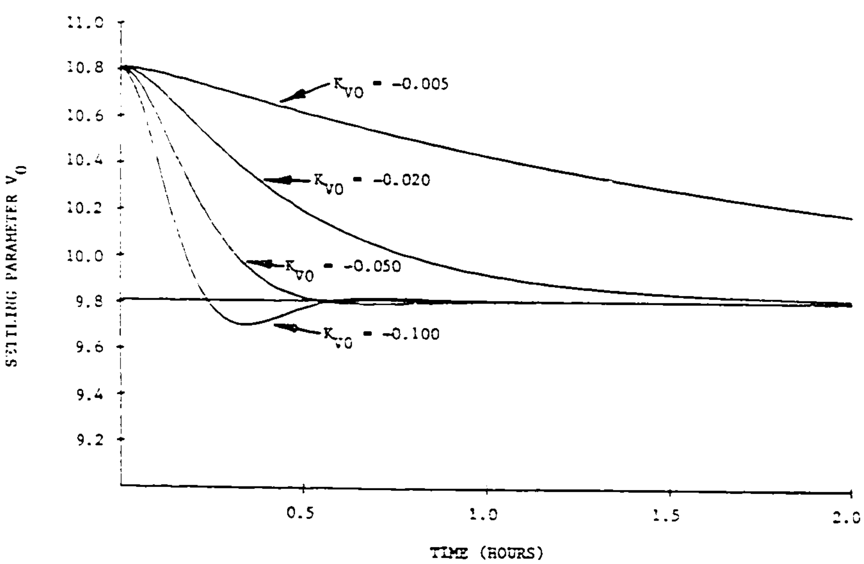

FIGURE 7.7 EFFECT OF ESTIMATOR COEFFICIENTS ON RATE OF CONVERGENCE OF SETTLING PARAMETER V_0 (K_{est}=10.0)

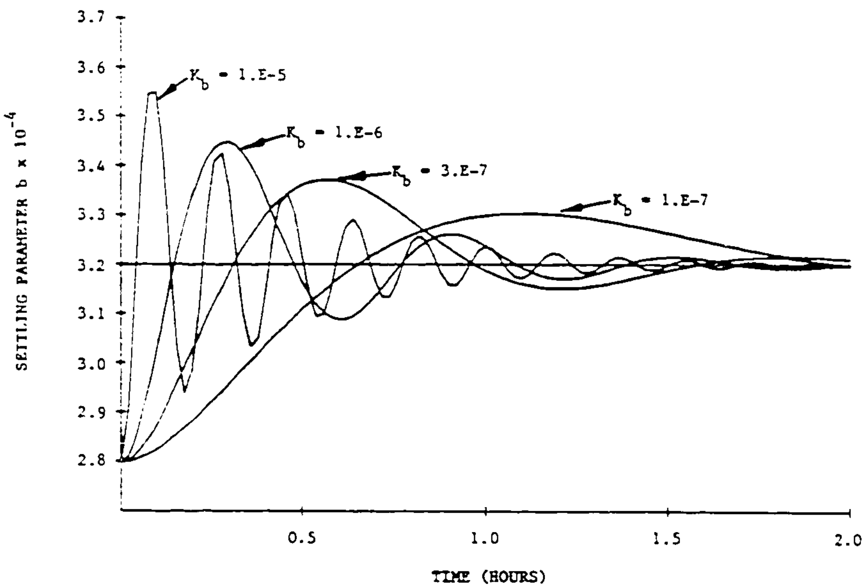

FIGURE 7.8 EFFECT OF ESTIMATOR COEFFICIENTS ON RATE OF CONVERGENCE OF SETTLING PARAMETER b (K_{est}=1.0)

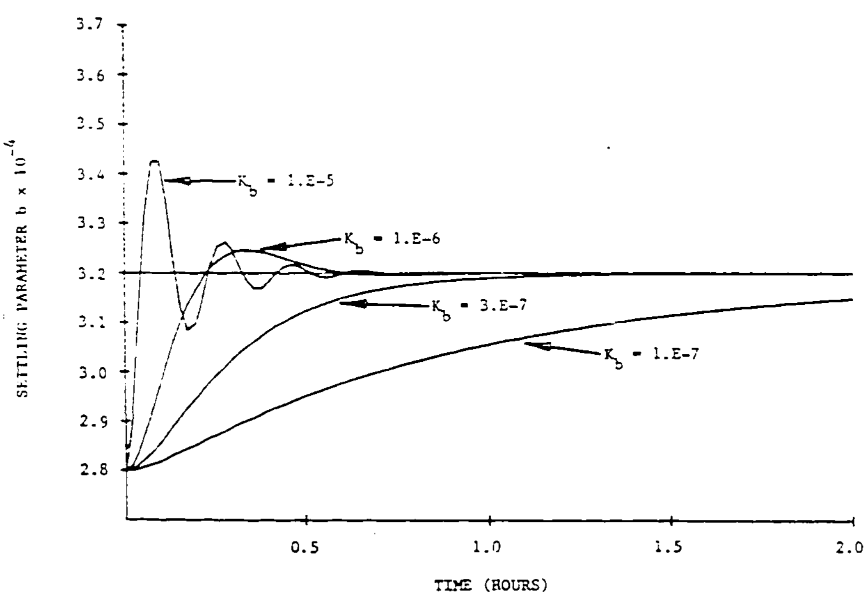

FIGURE 7.9 EFFECT OF ESTIMATOR COEFFICIENTS ON RATE OF CONVERGENCE OF SETTLING PARAMETER b (K_{est}=10.0)

Attempts to estimate both settling parameters simultaneously were not as successful as the single parameter estimations. This is understandable since the actual value which can be estimated from changes in solids concentration is the settling velocity itself, not the equation parameters. For every single settling velocity, there are an infinite number of V_0 and b pairs which can describe the observed phenomenon. Attempts to incorporate more than one model element into the estimators (Eqs. [7.13] and [7.14]) were also not successful.

Summary of Recursive Estimation Results

The objectives of using the recursive estimation procedure were to estimate the time-varying model parameters and to simplify the models necessary to adequately describe the activated sludge settler. This technique would ultimately be used for process control. It was shown that Olsson's (88) technique could be used to modify the state equations and provide a better estimator by incorporating process measurements. Equations for estimating the model parameters were also developed. Actual field data were used to develop the estimator coefficients for the clarification model while simulations were utilized to get coefficient values in the thickening model.

VIII. ENGINEERING SIGNIFICANCE

General

It is often noted that the engineering significance of any research depends upon what new knowledge is gained and the applicability of that knowledge to existing and anticipated problems. This section will discuss some of the new knowledge gained from this investigation and applications of the models developed and field verified in the course of this study. The majority of this work is oriented to the operation of the activated sludge process with special emphasis on the solid-liquid separation phase. However, there are also several design implications which will be discussed as well.

In section IV, four different models were developed. It was demonstrated that the clarification function of the settler was primarily a hydraulic phenomenon with a response time of minutes. Quite logically then, applications of the clarification model must be coupled with the hydraulic model which is capable of predicting the minute-to-minute flow changes. The settler thickening model, on the other hand, has a response time of many minutes to hours. Therefore, it is reasonable to couple it with the reactor mixing model which does not consider high frequency flow transients.

Applications of Hydraulic Control

The purpose of the hydraulic model is to predict hydraulic transients through the treatment plant induced by the operation of the influent and recycle pumps. These minute-to-minute predictions of flow can be coupled with the clarification model to predict the effects of hydraulic disturbances.

Influent Flow

The hydraulic model can be used to demonstrate the limited hydraulic buffering capacity of most wastewater treatment plants to influent flow disturbances. This buffering capacity is a nonlinear function of both magnitude and frequency of the disturbance. For the example shown, the influent flow was varied sinusoidally between 2 and 6 MGD (7.6 to 22.7 Ml/day) with several different frequencies. Recycle flow was maintained at a constant 2 MGD (7.6 Ml/day). The magnitude and phase angle of the plant outflow are shown in Figure 8.1. The results of Bryant (14) are also shown in Figure 8.1. The plant he modeled had considerably more hydraulic dampening capacity than the Sagemont plant.

Examination of Figure 8.1 reveals the limited buffering capacity of the Sagemont plant. For frequencies less than one cycle per hour, there is minimal attenuation. Even at four cycles/hour, the output magnitude is still 44 percent of the input value. This four cycles/hour example is typical of field conditions and corresponds approximately to a pump which is on for 7-1/2 minutes and off for 7-1/2 minutes. The consequence of this property is that disturbances in influent flow are propagated through the plant with little attenuation. The effects of these hydraulic transients have been demonstrated to degrade the effluent quality and should be avoided whenever possible.

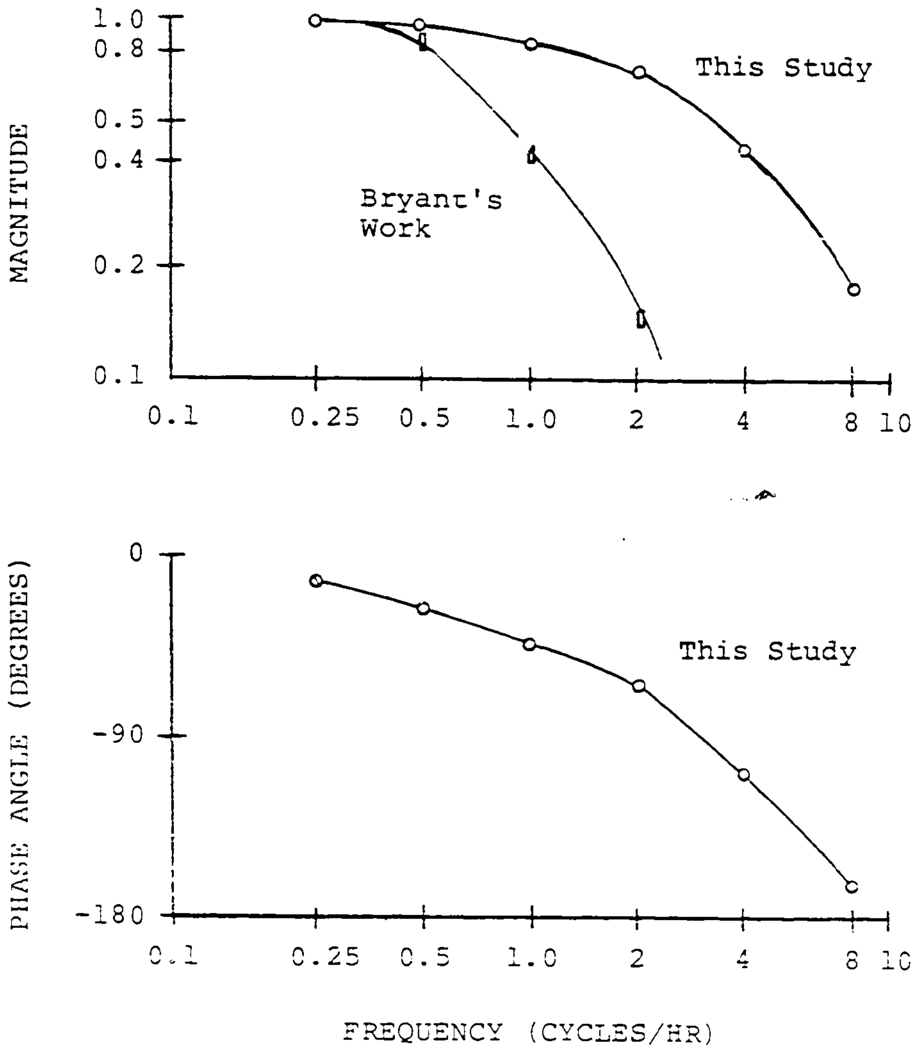

FIGURE 8.1 HYDRAULIC RESPONSE OF TREATMENT PLANT

There are a number of ways to dampen hydraulic transients including flow equalization, variable volume reactors, large head losses through the treatment plant, and improved pumping control.

Flow equalization is a design method which usually requires a large capital investment. Additionally, it usually entails additional pumps and aeration equipment and their associated costs. Due to the large diurnal flow variations experienced by municipal wastewater treatment plants, complete flow equalization is seldom economical. One possible exception to this rule-of-thumb is the small package treatment plant.

Small plants represent the majority of the number of plants and treat a significant (though much smaller) portion of the total wastewater flow. A survey of 1,612 plants in the state of Texas shows that 85 percent of the plants have a rated capacity of one MGD (3.85 Ml/day) or less and treat 15 percent of the aggregate flow. Small plants suffer from a number of ailments specific to their size and have gained a bad reputation with many regulatory agencies. A major problem is that of hydraulic surges caused by oversized pump stations. Due to their small capacity, this problem can be economically corrected with the addition of a small surge tank of 10-20 minutes detention time.

It is also possible to equalize flow by using variable volume reactors. Speece and LaGrega (104) discuss this alternative and some of its operational problems including the meeting of mixing and oxygenation requirements with a varying liquid level.

The use of large head losses in the plant to dampen hydraulic surges is often effectively practiced in hydraulically underloaded plants by partially closing gates and valves. Few plants, however, have sufficient freeboard for this practice.

There are several potential applications for the proposed influent pump control strategies. For overloaded plants not meeting their permit limits, a large portion of the effluent load is usually caused by effluent solids. Implementation of the pump control strategy could decrease the effluent solids and improve the effluent quality. For plants which usually require filters to consistently meet their permit requirements, addition of the pumping strategy might add equivalent consistency without the use of filters.

Both of these applications could be implemented with a relatively small investment of equipment (e.g., control valve, flow meter, and controller) and could possibly delay costly plant expansions such as the addition of reactors, settlers, and/or filters. Additionally, since there is little hardware associated with the pump strategy, it could be implemented much faster than a plant expansion.

For underloaded plants easily meeting their permit limits, the pump control strategy minimizing effluent solids shows little benefit. In fact, its use costs extra money since additional solids must be processed. Solids processing and disposal is very costly with costs of up to $300 per dry ton not being unusual. For these plants it is possible to restructure the minimization routine to maintain the effluent solids just under the permit limit. While this strategy would increase the effluent solids, this increase would still be within the permit limits and therefore legally acceptable. For example, at a 100

MGD (385 M l/day) plant an increase in one mg/l effluent solids would decrease the waste solids by 834 pounds (380 kg) per day. At $300 per dry ton, this represents a savings of $46,000 annually.

Recycle Flow

Simulations varying the recycle flow rate revealed an interesting phenomenon. The two simulations included a step increase (2 to 6 MGD)and a step decrease (6 to 2 MGD) in recycle flow rate while maintaining a constant influent flow rate (4 MGD). The step decrease in recycle flow resulted in an almost immediate increase in overflow rate in the settler. This sudden increase died out with a time constant of about 5-1/2 minutes. Likewise, the step increase in recycle flow resulted in a sudden decrease in overflow rate with a similar recovery. Simulation results are shown in Figures 8.2 and 8.3. The settler parameters of section VI were used in the simulations.

The phenomenon in Figure 8.2 can be explained by noting that the flow coming into the settler is (F_0+Fr). The flow going over the weirs is (F_0+Fr-Fr) or (F_0). If the recycle flow is suddenly decreased from (Fr_1) to (Fr_2), the flow going into the clarifier is still (F_0+Fr_1) while the overflow is (F_0+Fr_1-Fr_2). This last quantity is greater than (F_0) momentarily but decreases back to (F_0) as the effects of decreasing the recycle flow propagate through the plant. This increase in overflow rate is also accompanied by an increase in effluent suspended solids (21 to 27 mg/l) and a general deterioration of effluent quality. Similar reasoning can be used to analyze the step increase in recycle flow.

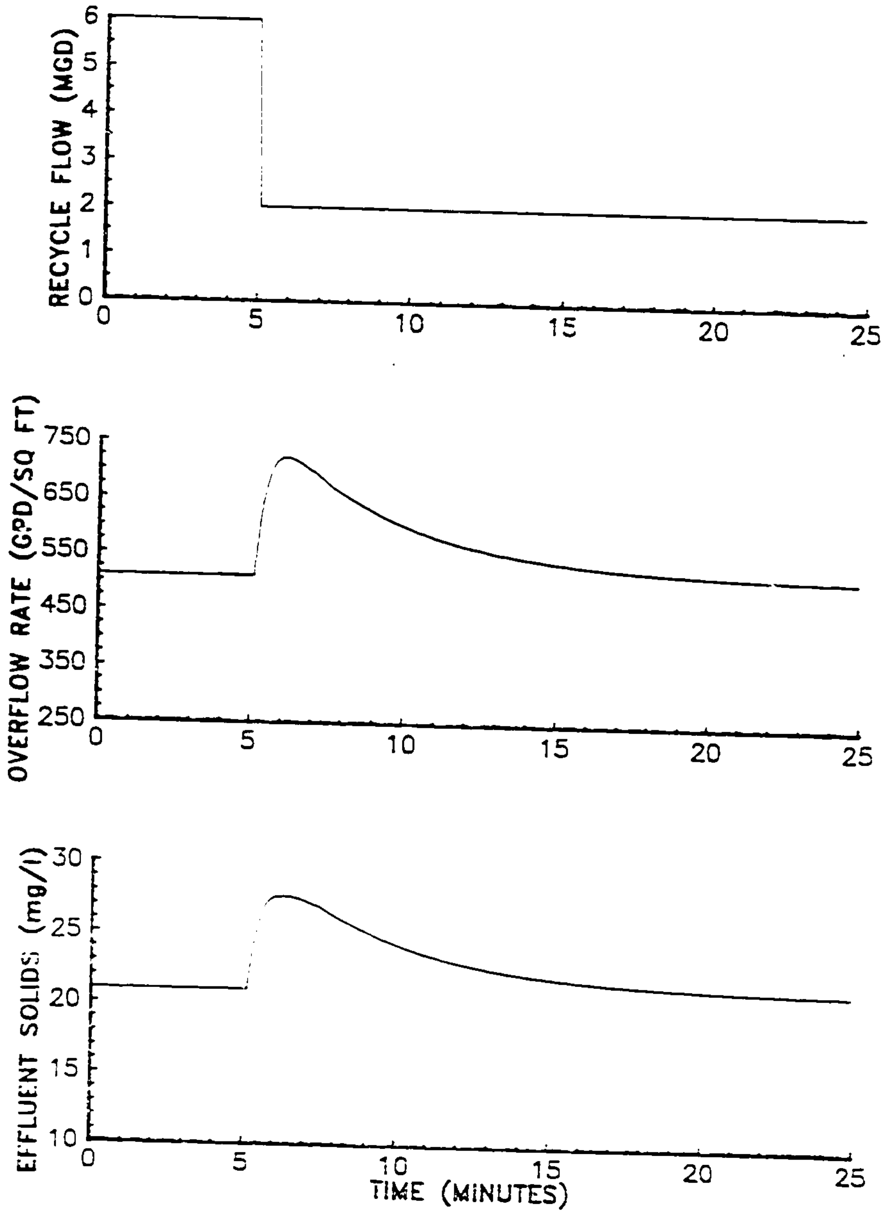

FIGURE 8.2 SETTLER RESPONSE TO STEP DECREASE IN RECYCLE FLOW RATE

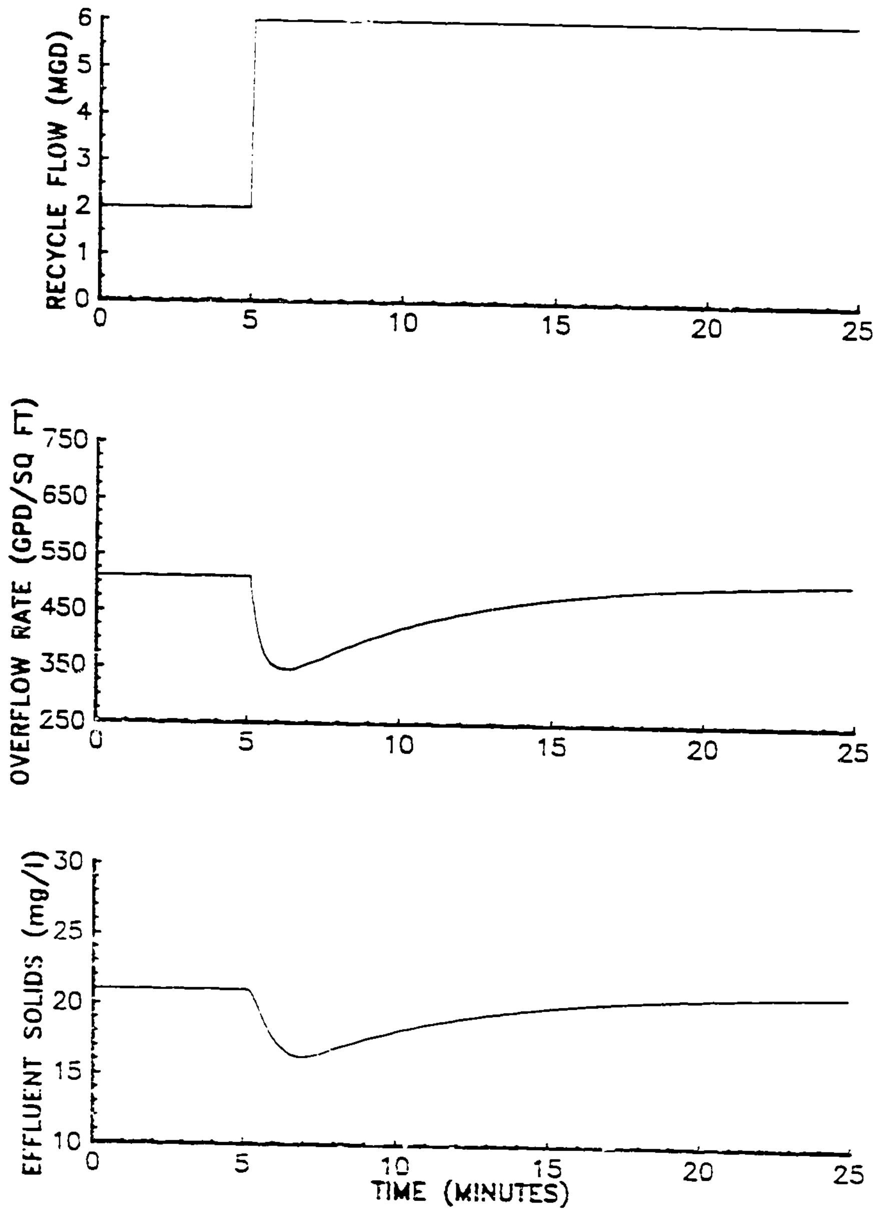

FIGURE 8.3 SETTLER RESPONSE TO STEP INCREASE IN RECYCLE FLOW RATE

As a result of these simulations it can be concluded that sudden decreases in recycle flow rate increase the overflow rate temporarily and degrade the effluent quality. Likewise, sudden increases in recycle flow rate decrease the overflow rate temporarily and may actually upgrade the effluent quality a small amount. This conclusion that decreases in recycle flow are more detrimental than increases is generally counter-intuitive and not discussed in manuals on the operation of the activated sludge process.

Applications of the Solids and Thickening Models

The purpose of the reactor mixing model is to predict the suspended solids concentrations in the various reactor under different hydraulic forcings. When coupled with the sludge thickening model, it can be used to predict the transfer of sludge between the reactors and the settler. Additionally, it can be used to predict the distribution of solids between the reactors when feed forward flow (step feed) is used. An example comparing experimental and simulation results was given in section VI.

This real-time knowledge of the solids distribution in the reactors is necessary information for many control strategies such as the SCOUR strategies previously discussed. This knowledge can be obtained in real-time by only two methods; on-line instrumentation or computer simulation. Laboratory methods for suspended solids (nonfiltrable residue) take approximately 1-1/2 hours when run in accordance with Standard Methods (107) and are thus not acceptable for on-line control purposes. The use of the solids model minimizes the use of on-line instruments to adjust the inputs to the reactor system. For multi-reactor configurations, this minimization of instruments can represent a significant savings.

Real-time knowledge of the settler underflow concentration and the amount of solids stored in the settler is also valuable information. Many of the control strategies developed elsewhere (2, 135) require a knowledge of the recycle solids concentration. Again, the reasonable alternatives for obtaining these values are on-line instruments and computer simulation. Elimination of any instruments can be a significant savings both in initial and maintenance costs. These two models also allow the prediction of gross process failure by loss of the sludge blanket over the effluent weirs. Bernard (9) emphasizes the prevention of this as the first priority of any activated sludge control strategy. Torpey (120) demonstrated the use of step feed for this purpose.

Applications to Computer Control

The on-line instruments and computer monitoring system utilized during this investigation allowed the observation of the dynamic behavior of the solids in the activated sludge process. It was observed that some variables of interest such as MLSS concentration (see Figures 4.7 and 4.8) and sludge blanket height (see Figure 4.18) varied relatively slowly (many minutes to hours). It was also observed that other variables such as flow (see Figure 6.1), effluent suspended solids (see

Figures 4.6, 4.7, and 4.8), and return sludge concentration (see Figure 4.27) varied on a minute-to-minute basis.

The occurrance of rapid changes in solids concentr.cions in the activated sludge process is not well known. Most engineers still think of the process as changing very slowly because of the long hydraulic residence times in the basins. In fact, this belief is often quoted as a major reason for manual rather than automatic operation of treatment plants. This investigation has shown that a number of variables change too rapidly to be adequately controlled by manual operation. These rapid variations also have an effect on downstream units and tend to propagate through the treatment train. This is justification for the use of computer control in wastewater treatment plants.

Applications to Sampling

The rapid changes observed in some of the variables of interest have widespread significance to sampling for performance monitoring and control. This investigation has shown instances in which the return sludge concentration can vary by up to 6,000 mg/l over a five minute interval even when the flows are constant. The usefulness of a single grab sample from this stream is therefore questionable. However, this type of sampling is often used to control the return sludge flow rate and the sludge wastage rate. A more reasonable approach would be to average the sample over some time period approximating the frequency of the disturbance. This could be done by physically compositing a sample or filtering the signal from the on-line instrument.

Design of Clarifier Outlets

Almost without exception, the outlets of clarifiers are triangular weirs. The "correct" location of these weirs, however, is still a subject of debate. This was evidenced recently when two article in the same issue of the Journal of the Water Pollution Control Federation suggested different locations for effluent weirs. Parker (92) suggested moving the weirs toward the center of the settler while Stukenberg, et al. (114) suggested moving them to the periphery. Most sources say that triangular weirs are used so that the flow distribution along the outlet will be relatively constant even if the weirs are not level. Other advantages of triangular weirs include the ability to handle a wide range of flows, ease of cleaning, and a no-clog configuration.

It can be easily shown, however, that triangular weirs could be a bad choice for even flow distribution when weirs are not level (7). This results from the fact that flow is proportional to the head to the 2.5 power for triangular weirs. An increase in head (from weir plates at different elevations) of 50 percent will increase the flow by 176 percent. For rectangular weirs, the flow is proportional to the 1.5 power. In this case a 50 percent increase in head results in an 84 percent increase in flow. Carrying this idea one step further, the flow from a submerged orifice is proportional to head to the 1/2 power. For the orifice, a 50 percent increase in head results in only a 22 percent increase in flow. Thus, to compensate for differences in head, a submerged orifice will maintain a more even flow distribution along the entire length of the weirs than either triangular or rectangular weirs.

The submerged orifice also has the advantage of drawing liquid from below the water surface, avoiding floating debris. The major disadvantages of submerged weirs are their limited flow range and the possibility of clogging. Lutge (76) has reported the successful use of submerged orifices on twelve large, rectangular settlers in Seattle.

IX. CONCLUSIONS

The following conclusions were made as a result of the experimental investigations and computer simulations conducted during this study. It should be noted that some of these conclusions are directly applicable only to the wastewater plant investigated which was lightly loaded and had a sludge with good settling and clarification characteristics.

1. A reactor mixing model based on completely mixed reactors in series can be used to predict MLSS concentrations in the activated sludge aeration basins. The reactors have a considerable concentration buffering capacity which exhibits a time constant on the order of minutes to hours.

2. A hydraulic model based upon mass balances and well known flow equations can be used to predict the propagation of hydraulic disturbances through the treatment plant. These disturbances are caused by the operation of influent and recycle pumps. The Sagemont plant had little hydraulic buffering capacity and exhibited a time constant on the order of five minutes.

3. Sudden changes in the recycle flow rate can affect the settler by temporarily changing the overflow rate. Increases of recycle flow will momentarily decrease the overflow rate while decreases in recycle flow will increase the overflow rate. These changes are relatively short lived due to the small hydraulic buffering capacity of the plant and die out within ten to twenty minutes.

4. Underflow rates of up to 575 gpd/sq ft (1.0 m/hr) have little effect on the effluent suspended solids concentration except for hydraulic transients induced by sudden changes. There is qualitative evidence from other plants, however, that suggest underflow rates above 1000 gpd/sq ft (1.7 m/hr) may lead to a degradation of effluent quality due to increased turbulence in the settler. It was not possible to conduct tests in this range at the plant under study.

5. Qualitatively it was observed that effluent suspended solids increase with increasing MLSS concentrations. It was not possible to evaluate a function for this relationship since only two solids levels were studied.

6. It was found that the sludge blanket level did not greatly affect the effluent suspended solids concentration at heights of up to 4.3 feet (1.3 m) below the water surface and overflow rates up to 1100 gpd/sq ft (1.9 m/hr). At some level above this, however, the blanket solids will be carried over the weirs resulting in a large discharge of solids. This gross process failure is the result of a thickener failure, not a clarification failure.

7. It was observed that hydraulic transients have an immediate effect on effluent suspended solids concentration which can persist longer than the flow disturbance. Thus, both influent flow rate and pattern have an effect on effluent suspended solids. This fact suggests that the on/off pumping strategy employed at most sewage lift stations may be detrimental to effluent quality.

8. A dynamic sludge thickening model based on solids flux concepts can be used to predict underflow solids concentrations when a sludge blanket exists. An empirical model based on the same assumptions but which accounts for the geometry of the conical section of the settler can be used to predict underflow solids concentrations for under-loaded, over-loaded, and transition conditions.

9. The recycle solids concentration exhibits periodic (13 minutes) peaks corresponding to the passage of the sludge rake over the sludge hopper. Concentration changes may be up to 5,000 or 6,000 mg/l over this short time period. Because of this property, manual or automatic control systems which use the recycle solids concentration should use a time averaged value. Simple grab samples may lead to large errors.

10. The on/off operation of influent pumps results in the loss of more effluent suspended solids due to flow transients than the use of a variable flow controller utilizing a constant level algorithm. The optimization routine for controlling influent flow rate developed in this investigation has the potential for producing a higher quality effluent than either of the other two control methods by utilizing the lift station wet well volume in an optimal manner.

11. Recursive estimation can be utilized to estimate the time varying parameters of the models as well as variables which cannot be easily measured.

X. RECOMMENDATIONS FOR FUTURE WORK

Based upon the results of the experimental investigation and computer simulations conducted during this study, the following recommendations for further work are made:

1. More theoretical and experimental work is needed to establish the mechanisms by which clarification occurs in secondary sedimentation.

2. A more quantitative relationship between settling characteristics and conditions in the biological reactors is needed. This relationship is needed to couple control strategies based on biological needs in the reactors and physical requirements in the settler.

3. The adsorption and desorption of primary particles present in the influent should be investigated and modeled in order to further refine the use of feed forward flow as a control strategy.

4. A thorough economic evaluation should be performed on the influent flow control strategy to delineate under what conditions it would be the best alternative to upgrade the performance of a wastewater treatment plant.

5. More theoretical and experimental work is needed to refine the state/parameter estimation technique. In particular, a reduction in the number of required measurements is desirable. The use of sludge blanket level may be helpful for estimating sludge settling parameters and should be investigated.

6. There is a need to blend models such as those developed in this research with stochastic models to account for the random errors which cannot be accounted for by deterministic models. This blending of modeling techniques could lead to better prediction of plant performance and better control.

APPENDIX A

NOMENCLATURE

Symbol	Description
a	Settling velocity equation parameter
$a(i)$	Fourier series coefficient
a_1	Model parameter
A	Settler surface area (L^2)
AREA	Settler surface area (L^2)
b	Settling velocity equation parameter
$b(i)$	Fourier series coefficient
BOD_5	Five day biochemical oxygen demand (M/L^3)
c_0, c_1, c_2	Concentrations, usually (M/L^3)
$c(i)$	Polynomial coefficients
C	Normalized concentration
C_W	Weir coefficient
d_l	density of liquid (M/L^3)
D	Axial dispersion coefficient (L^2/T)
DO	Dissolved oxygen concentration (M/L^3)
DT	Detention time (T)
(D/uL)	Vessel dispersion number
E_p	Penalty error term due to constraint violations
E_T	Total error
E_x	Error term due to discharge of solids
$\hat{f}$	Predicted flow (L^3/T)
$\bar{f}$	Historical average flow (L^3/T)

NOMENCLATURE (continued)

Symbol	Description
fc	Flow correction term (L^3/T)
F	Resistivity ($M/L^3/T$)
F_{air}	Air flow rate (L^3/T)
Fe	Effiuent flow rate (L^3/T)
F_{in}	Flow in (L^3/T)
F_{max}	Maximum desired outflow (L^3/T)
F_{out}	Flow out (L^3/T)
Fr	Recycle flow rate (L^3/T)
FTU	Formazone turbidity unit
F_0	Influent flow rate (L^3/T)
Fw	Waste flow rate (L^3/T)
F/M	Food to microorganisms ratio ($M\text{-}BOD_5/M\text{-}MLSS$)
g	Acceleration of gravity (L/T^2)
Gs	Flux due to settling velocity ($M/L^2/T$)
h	Depth of liquid over weir (L)
H	Wet well water height (L)
H_{max}	Maximum desired wet well height (L)
H_{min}	Minimum desired wet well height (L)
Hz	Hertz, cycles per second (T^{-1})
ISV	Initial settling velocity (L/T)
K	Unit conversion factor
K	Headloss coefficient
K	Constant correction matrix
K_D	Estimator coefficient

NOMENCLATURE (continued)

Symbol	Description
Kd	Organism decay rate coefficient (T^{-1})
K_{est}	Correction matrix coefficients, constant
K_N	Model parameter
K_{V0}	Estimator coefficient
K_1, K_2, K_3	Model parameters
L	Length of vessel, length of weir (L)
M	Solids transported through settler (M)
$\bar{M}$	Solids transported through settler at a constant velocity (M)
MGD	Million gallons per day (L^3/T)
MIN	Minimum function
MLSS	Mixed liquor suspended solids (M/L^3)
M l/day	Million liters per day (L^3/T)
N	Number of tanks in series
p	Liquid phase pressure ($M/L/T^2$)
P	Vector of model parameters
PI	Pi, 3.14159···
Q	Liquid flow rate (L^3/T)
$\hat{Q}$	Operating point for flow rate (L^3/T)
r, R	Recycle ratio (Fr/F_0)
R	Hydraulic radius (L)
R	Hydraulic resistance ($L/L^3/T$)
R	Optimization evaluation response
RAS	Return activated sludge
S	Slope of the energy line

NOMENCLATURE (continued)

Symbol	Description
SCOUR	Specific Carbonaeous Oxygen Uptake Rate (M-O_2/T/M-X)
SRT	Solids retention time, Sludge Age (T)
SS	Suspended solids concentration (M/L^3)
SSV	Stirred specific volume (L^3/M)
SSVI	Standard sludge volume index (L^3/M)
SVI	Sludge volume index (L^3/M)
SWD	Side water depth (L)
t, T	Time (T)
T	Planning horizon (T)
$\bar{t}$	Centroid of distribution, mean (T)
u	Velocity of flow (L/T)
U	Net downward velocity due to sludge removal (L/T)
U(t)	Step function = 0 if $t<0$ and 1 if $t>0$
V	Volume (L^3)
V	Instantaneous velocity (L/T)
$\bar{V}$	Time averaged velocity (L/T)
V_s	Settling velocity (L/T)
Vs	Volume of sludge in settler (L^3)
V_T	Velocity characteristic of turbulence level (L/T)
V_w	Velocity of displaced fluid (L/T)
V_0	Stoke's settling velocity for a single discrete particle (L/T)
V_{30}	MLSS 30-minute settled volume (as a fraction)
y	Process measurements
Y	Yield coefficient (M-X/M-BOD_5)

NOMENCLATURE (continued)

Symbol	Description
X	Solids concentration (M/L^3)
X	Spatial distance (L)
X	Matrix of measurements, observations, etc.
X^T	Transpose of matrix X
$\hat{X}$	Model states
Xe	Effluent suspended solids (M/L^3)
$(Xe)_{ss}$	Steady state effluent solids concentration (M/L^3)
X_M	Mass of solids
X_{sb}	Solids concentration in top element of sludge blanket zone (M/L^3)
$\hat{X}_{sb}$	Estimated solids concentration in top element of sludge blanket zone (M/L^3)
Xw	Solids concentration of waste stream (M/L^3)
X_{1-19}	Solids concentration in model elements (M/L^3)
Z	Vertical distance (L)
ΔA	Differential area (L^2)
ΔV	Differential volume (L^3)
θ	Normalized time, tu/L
σ	Effective interparticle pressure ($M/L/T^2$)
σ^2	Variance
π	Pi, 3.14159···
ρ_l	Bulk density of liquid (M/L^3)
ρ_s	Bulk density of solids (M/L^3)

APPENDIX B

COMPUTER PROGRAM LISTINGS

```
PROGRAM THICK10

"Dynamic Model of Settler Thickening Dynamics"
"Recursive Algorithm for State/Parameter Estimation"
"Originally written in the CSMP Language: February 1984"
"Rewritten in the ACSL Language: October 29, 1984"
"ADVANCED CONTINUOUS SIMULATION LANGUAGE"
"Mitchell and Gauthier, Assoc., Inc."
"290 Baker Avenue"
"Concord, Mass.  01742"

"Robert D. Hill"
"Visiting Assistant Professor"
"McMaster University"
"Hamilton, Ontario, Canada"

INITIAL

"Dimension Variables"
   INTEGER LU
   REAL V(19), A(19), XI(19), X(19), DX(19)
   REAL XHI(19), XH(19), DXH(19), KEST(19)
   "V(i)  = Settler Slice Volumes, cu m"
   "A(i)  = Settler Slice Surface Areas, sq m"
   "X(i)  = Settler Slice Solids Concentrations, mg/l"
   "XI(i) = Settler Slice Initial Conditions, mg/l"
   "DX(i) = Derivative of X(i), mg/l/hr"
   "XH(i)  = Estimated Solids Concentration, mg/l"
   "XHI(i) = Initial Conditions of XH(i), mg/l"
   "DXH(i) = Derivative of XH(i), mg/l/hr"

" Define Simulation Time Constants, hrs"
   CINTERVAL CINT=0.02
   ALGORITHM IALG=5
   MAXTERVAL MAXT=0.01
   MINTERVAL MINT=1.E-6
   NSTEPS NSTP=10
   CONSTANT TSTOP=4.0

"Define Physical Constants"
   CONSTANT PI=3.14159

"Define Initial Values of Other Variables"
   CONSTANT NTIME=0.0, DTIME=15.0
   CONSTANT MLSS=4000.0, XE=0.0
```

```
"Read default values from data files"
   CALL INIT01 (V,A,XI,V0,B,XHI,V0H,BH,KEST,KV0,KB)
   V0HI=V0H
   BHI=BH

"Calculate Recycle Flow Rate"
   PROCEDURAL (FR=F0,T)
      FR = 320.0
   END $ "OF PROCEDURAL"

"Write Parameter Values to OUTPUT and PRINT files"
      DO 300 LU=6,9,3
      WRITE (LU,100) V0,B
100.. FORMAT (/1X,'V0 = ',F16.8,/1X,'B  = ',F16.8)
      WRITE (LU,200) V0H,BH
200.. FORMAT (/1X,'V0H = ',F16.8,/1X,'BH = ',F16.8,///)
300.. CONTINUE


END $ "OF INITIAL"

DYNAMIC
DERIVATIVE


"Calculate Solids Derivatives"
   CALL XDER (X,DX,V,A,F0,FR,MLSS,XE,V0,B)
   CALL XHDER (X,XH,DXH,V,A,F0,FR,MLSS,XE,V0H,BH,KEST)
   CALL PARDER (X,XH,DV0H,DBH,KV0,KB)

"Integrate Equations"
   X = INTVC (DX,XI)
   XH = INTVC (DXH,XHI)
   V0H = INTEG (DV0H,V0HI)
   BH = INTEG(DBH,BHI)

END $ "OF DERIVATIVE"


"Calculate Influent Flow Rate"
   PROCEDURAL (F0=T)
      F0=640.0
"      IF ((T.GE.1.0).AND.(T.LT.7.0)) F0=1280.0"
      "F0 = 950.0 + 315.0*SIN(PI*T/2.0)"
   END $ "OF PROCEDURAL"

   CALL OUT1 (T,X,XH,V0H,BH)

"End Condition"
TERMT (T.GE.TSTOP)

END $ "OF DYNAMIC"
END $ "OF PROGRAM
```

```
      SUBROUTINE INIT01 (V,A,XI,V0,B,XHI,V0H,BH,KEST,KV0,KB)

      INTEGER  I
      REAL     V(19), A(19), XI(19), V0, B
      REAL     XHI(19), V0H, BH, KEST(19), KV0, KB

      OPEN (UNIT=10, FILE='THICK10.INP', STATUS='OLD')
      OPEN (UNIT=11, FILE='THICK10.OUT', STATUS='NEW')

      READ (10,*) V0, B

      READ (10,*) (V(I),I=1,5)
      READ (10,*) (V(I),I=6,10)
      READ (10,*) (V(I),I=11,15)
      READ (10,*) (V(I),I=16,19)

      READ (10,*) (A(I),I=1,5)
      READ (10,*) (A(I),I=6,10)
      READ (10,*) (A(I),I=11,15)
      READ (10,*) (A(I),I=16,19)

      READ (10,*) (XI(I),I=1,5)
      READ (10,*) (XI(I),I=6,10)
      READ (10,*) (XI(I),I=11,15)
      READ (10,*) (XI(I),I=16,19)

      READ (10,*) V0H,BH

      READ (10,*) (XHI(I),I=1,5)
      READ (10,*) (XHI(I),I=6,10)
      READ (10,*) (XHI(I),I=11,15)
      READ (10,*) (XHI(I),I=16,19)

      READ (10,*) (KEST(I),I=1,5)
      READ (10,*) (KEST(I),I=6,10)
      READ (10,*) (KEST(I),I=11,15)
      READ (10,*) (KEST(I),I=16,19)

      READ (10,*) KV0,KB

200   CLOSE (UNIT=10)

      RETURN
      END


      SUBROUTINE XDER (X,DX,V,A,F0,FR,MLSS,XE,V0,B)

      INTEGER I
      REAL    X(19),DX(19),V(19),A(19),F0,FR,MLSS,XE,V0,B
      REAL    FLUXIN,VS(19),GS(19)

      FLUXIN = (F0+FR)*MLSS - F0*XE
```

```
      DO 100 I=1,19
      VS(I)=V0*EXP(-B*X(I))
      IF (X(I).LT.2000.0) VS(I)=20.0
      GS(I)=X(I)*VS(I)
100   CONTINUE

      DX(1)=(FLUXIN-FR*X(1))/V(1) - A(1)*(AMIN1(GS(1),GS(2)))/V(1)

      DO 200 I=2,18
      TEMP = A(I-1)*AMIN1(GS(I-1),GS(I)) - A(I)*AMIN1(GS(I),GS(I+1))
      DX(I)=FR*(X(I-1)-X(I))/V(I) + TEMP/V(I)
200   CONTINUE

      DX(19)=FR*(X(18)-X(19))/V(19) + A(18)*GS(19)/V(19)

      RETURN
      END


      SUBROUTINE XHDER (X,XH,DXH,V,A,F0,FR,MLSS,XE,V0H,BH,KEST)

      INTEGER I
      REAL    X(19),DXH(19),V(19),A(19),F0,FR,MLSS,XE,V0H,BH
      REAL    XH(19), KEST(19)
      REAL    FLUXIN,VS(19),GS(19)

      FLUXIN = (F0+FR)*MLSS - F0*XE

      DO 100 I=1,19
      VS(I)=V0H*EXP(-BH*XH(I))
      IF (XH(I).LT.2000.0) VS(I)=20.0
      GS(I)=XH(I)*VS(I)
100   CONTINUE

      DXH(1)=(FLUXIN-FR*XH(1))/V(1) - A(1)*(AMIN1(GS(1),GS(2)))/V(1) +
     1      KEST(1)*(X(1)-XH(1))

      DO 200 I=2,18
      TEMP = A(I-1)*AMIN1(GS(I-1),GS(I)) - A(I)*AMIN1(GS(I),GS(I+1))
      DXH(I)=FR*(XH(I-1)-XH(I))/V(I) + TEMP/V(I) + KEST(I)*(X(I)-XH(I))
200   CONTINUE

      DXH(19) = FR*(XH(18)-XH(19))/V(19) + A(18)*GS(19)/V(19) +
     1       KEST(19)*(X(19)-XH(19))

      RETURN
      END
```

```
      SUBROUTINE PARDER (X,XH,DVOH,DBH,KVO,KB)

      REAL X(19), XH(19), DVOH, DBH, KVO, KB

      DVOH = 0.0
      DBH = 0.0

      DO 100 I=1,19
      IF (X(I).LT.2000.0) GOTO 100
      DVOH = DVOH + KVO*(X(I)-XH(I))
C      DBH  = DBH  + KB *(X(I)-XH(I))
      GOTO 1000
100   CONTINUE

1000  RETURN
      END


      SUBROUTINE OUT1 (T,X,XH,VOH,BH)

      INTEGER I
      REAL T, X(19), XH(19), VOH, BH

      WRITE (11,*) T,VOH,BH

      RETURN
      END
```

```
PROGRAM OPTIMAL

"Dynamic Model of Reactor Hydraulics"
"with Prediction of Effluent Suspended Solids"
"Optimizing Algorithm for Controlling Lift Station Output"
"Originally written in the CSMP Language: September 12, 1983"
"Rewritten in the ACSL Language: October 21, 1984"
"ADVANCED CONTINUOUS SIMULATION LANGUAGE"
"Mitchell and Gauthier, Assoc., Inc."
"290 Baker Avenue"
"Concord, Mass.  01742"

"Robert D. Hill"
"Visiting Assistant Professor"
"McMaster University"
"Hamilton, Ontario, Canada"

INITIAL

" Define Simulation Time Constants, min"
   CINTERVAL CINT=1.0
   ALGORITHM IALG=5
   MAXTERVAL MAXT=1.0
   MINTERVAL MINT=1.E-6
   NSTEPS NSTP=100
   CONSTANT TSTOP=1440.0

"Define Initial Conditions, ft"
   CONSTANT HBS0=2.3133E-3, H10=6.8617E-2, H1A0=6.8579E-2
   CONSTANT H20=6.8267E-2,  H30=6.8077E-2, H40=6.8039E-2
   CONSTANT HCF0=6.3526E-2, HR10=6.9898E-2, HR20=6.8928E-2
   CONSTANT HCL10=8.6773E-2, HCL20=8.6703E-2, HOUT0=0.94573
   CONSTANT HWW0=5.0, XE0=1.86

"Define Physical Constants"
   CONSTANT PI=3.14159, G=32.2

"Define Reactor Surface Areas, sq ft"
   CONSTANT ABARS=420.0, A1=3735.0, A1A=2470.0, A2=3300.0
   CONSTANT A3=1847.0, A4=4500.0, ACF=7834.0, ACL1=280.0
   CONSTANT ACL2=3120.0, AOUT=666.0, AR1=2180.0, AR2=923.0
   CONSTANT AWW=508.0

"Define Resistance Constants for Gates, Weirs, etc."
   CONSTANT NCF=665.0, NCL=350.0
   CONSTANT K1=0.5, K1A=0.5, K2=0.5, K4=1.5
   CONSTANT KCL1=0.5, KR1=1.0, KR2=0.5

"Define Recycle Flow Rate at a Constant 50%"
   CONSTANT R=0.5

"Define Other Constants"
   CONSTANT C1=5.0, C2=4.0, C3=-0.032
```

```
"Define Initial Values of Other Variables"
   CONSTANT NTIME=0.0, DTIME=15.0

"Read default values from data files"
"Put values in FORTRAN common areas"
   CALL INIT01

END $ "OF INITIAL"

DYNAMIC
DERIVATIVE

"Define Derivatives of all Reactor Water Levels"

"Wet Well Level"
DHWW=(FIN-FO)/AWW
HWW=INTEG(DHWW,HWW0)

"Bar Screen Area"
PROCEDURAL (FBS=HBS)
   IF (HBS.GT.0.0) FBS=60.0*3.33*(20.0-2.0*HBS)*HBS**1.5
   IF (HBS.LE.0.0) FBS=0.0
END $ "OF PROCEDURAL"
DHBS=(FO-FBS)/ABARS
HBS=INTEG(DHBS,HBS0)

"Reactor #1"
Z1=H1-H1A
H1PRT=H1+46.33
A=2.0*(3.0+7.0/12.0)*(H1PRT-39.5)
F1=SIGN(1.0,Z1)*60.0*A*SQRT(2.0*G/K1)*SQRT(ABS(Z1))
DH1=(FBS-F1+FR2)/A1
H1=INTEG(DH1,H10)

"Reactor #1A"
Z1A=H1A-H2
"H1APRT=H1A+46.33"
F1A=SIGN(1.0,Z1A)*60.0*2.0*9.0*SQRT(2.0*G/K1A)*SQRT(ABS(Z1A))
DH1A=(F1-F1A)/A1A
H1A=INTEG(DH1A,H1A0)

"Reactor #2"
Z2=H2-H3
"H2PRT=H2+46.33"
F2=SIGN(1.0,Z2)*60.0*2.0*12.25*SQRT(2.0*G/K2)*SQRT(ABS(Z2))
DH2=(F1A-F2-FW)/A2
H2=INTEG(DH2,H20)
```

```
"Reactors #3 and #4"
"H3=H4, Therefore: DH3=DH4"
Z4=H4-HCF
"H3PRT=H3+46.33"
"H4PRT=H4+46.33"
F4=SIGN(1.0,Z4)*60.0*9.62*SQRT(2.0*G/K4)*SQRT(ABS(Z4))
F3=(A3*F4 + A4*(F2+FFF))/(A3+A4)
DH3=(F2+FFF-F3)/A3
DH4=(F3-F4)/A4
H3=INTEG(DH3,H30)
H4=INTEG(DH4,H40)

"Clarifier"
ZCF=HCF
PROCEDURAL (FCF=NCF,ZCF)
   IF (ZCF.GT.0.0) FCF=60.0*NCF*2.5*ZCF**2.5
   IF (ZCF.LE.0.0) FCF=0.0
END $ "OF PROCEDURAL"
DHCF=(F4-FCF-FR)/ACF
HCF=INTEG(DHCF,HCF0)

"Return Sludge Channel #1"
ZR1=HR1-HR2
"ZR1PRT=HR1+46.33"
FR1=SIGN(1.0,ZR1)*60.0*12.57*SQRT(2.0*G/KR1)*SQRT(ABS(ZR1))
DHR1=(FR-FR1)/AR1
HR1=INTEG(DHR1,HR10)

"Return Sludge Channel #2"
ZR2=HR2-H1
"ZR2PRT=HR2+46.33"
FR2=SIGN(1.0,ZR2)*60.0*16.0*SQRT(2.0*G/KR2)*SQRT(ABS(ZR2))
DHR2=(FR1-FR2)/AR2
HR2=INTEG(DHR2,HR20)

"Inlet To Chlorine Basin"
ZCL1=HCL1-HCL2
"ZC1PRT=HCL1+44.0"
FCL1=SIGN(1.0,ZCL1)*60.0*2.0*9.0*SQRT(2.0*G/KCL1)*SQRT(ABS(ZCL1))
DHCL1=(FCF-FCL1)/ACL1
HCL1=INTEG(DHCL1,HCL10)

"Chlorine Basin"
ZCL2=HCL2
"ZC2PRT=HCL2+44.0"
FCL2=60.0*NCL*2.5*ZCL2**2.5
DHCL2=(FCL1-FCL2)/ACL2
HCL2=INTEG(DHCL2,HCL20)
```

```
"Chlorine Basin Outlet"
ZOUT=HOUT
"ZOTPRT=HOUT+41.77"
FOUT=60.0*2.5*ZOUT**2.5
DHOUT=(FCL2-FOUT)/AOUT
HOUT=INTEG(DHOUT,HOUT0)

"Effluent suspended solids"
XESS = C1 + C2*FCF/92.82776
DXE = C3*(XE-XESS)*60.0
XE=INTEG(DXE,XE0)
PROCEDURAL (XE=XE,XESS)
  IF (XESS.GT.XE) XE=XESS
END $ "OF PROCEDURAL"

"Accumulate effluent solids and flow"
XETOT=INTEG(FCF*XE,0.0)
FCFTOT=INTEG(FCF,0.0)

"Get 15 minute average effluent flow"
F15MIN=INTEG(FOUT,0.0)

"End condition"
TERMT (T.GE.TSTOP)

END $ "OF DERIVATIVE"


"Calculate average effluent solids concentration"
PROCEDURAL (XEAVG=XETOT,FCFTOT)
  XEAVG = 0.0
  IF (FCFTOT.GT.0.0) XEAVG=XETOT/FCFTOT
END $ "OF PROCEDURAL"

"Turn on whichever controller is desired."
"Comment out the others"

"Calculate Influent Flow From Lift Station"
"      CALL ONOFF (F0,HWW)"
"Pump Station Level Controller"
"      CALL LEVEL (F0,HWW,T,NTIME,DTIME)"
"Optimize Flow Rate From Wet Well"
       CALL OPTIM (F0,FIN,T,NTIME,DTIME,F15MIN,HWW,C1,C2,C3)

"Calculate Influent Flow Rate"
   PROCEDURAL (FIN=T)
      CALL INFLOW (FIN,T)
      FIN=FIN*92.82776
   END $ "OF PROCEDURAL"
```

```
"Calculate Recycle Flow Rate"
   PROCEDURAL (FR=F0,R)
      FR=F0*R
      FR=AMIN1(465.0,FR)
   END $ "OF PROCEDURAL"

"Calculate Feed Forward (Step-Feed) Flow Rate"
   PROCEDURAL (FFF=T)
      FFF=0.0
   END $ "OF PROCEDURAL"

"Calculate Waste Flow Rate"
   PROCEDURAL (FW=T)
      FW=0.0
   END $ "OF PROCEDURAL"


END $ "OF DYNAMIC"

END $ "OF PROGRAM


      SUBROUTINE INIT01

      INTEGER  I, J
      REAL     F(1560)
      COMMON /RTDATA/ F
      DATA J /0/

      OPEN (UNIT=10, FILE='SEPT0183.1MN', STATUS='OLD')

100   READ (10,*,ERR=200,END=200) (F(6*J+I),I=1,6)
      J=J+1
      IF ((6*J+6).LE.1560) GOTO 100

200   CLOSE (UNIT=10)
      RETURN
      END


      SUBROUTINE INFLOW (FIN,T)

      INTEGER  I
      REAL     FIN, T, F(1560)

      COMMON /RTDATA/ F

      I = 1+INT(T)
      I = MIN0 (1560,I)
      FIN=F(I)

      RETURN
      END
```

```
      SUBROUTINE ONOFF (F0,HWW)

      REAL F0,HWW

      IF (HWW.LT.4.0) F0=0.0
      IF ((HWW.GE.9.0).AND.(F0.LT.465.0)) F0=465.0
      IF ((HWW.GE.11.0).AND.(F0.LT.930.0)) F0=930.0
      IF (HWW.GE.13.0) F0=1395.0

      RETURN
      END


      SUBROUTINE LEVEL (F0,HWW,T,NTIME,DTIME)

      REAL F0,HWW,T,NTIME,DTIME

      IF (T.LT.NTIME) GOTO 1000

      IF (HWW.LT.4.0) F0=0
      IF ((HWW.GE.4.0).AND.(HWW.LT.9.0).AND.(F0.GT.0.0)) F0=232.5

      IF ((HWW.GE.9.0).AND.(HWW.LT.11.0)) THEN
         IF (F0.LT.697.5)  F0 = 232.5 + 232.5*(HWW-6.0)
         IF (F0.GT.465.0)  F0 = 697.5
      ENDIF

      IF ((HWW.GE.11.0).AND.(HWW.LT.13.0)) THEN
         IF (F0.LT.1395.0)  F0 = 697.5 + 232.5*(HWW-7.0)
         IF (F0.GT.930.0)   F0 = 1395.0
      ENDIF

      IF (HWW.GE.13.0) F0=1395.0

      NTIME = NTIME + DTIME

1000  CONTINUE
      RETURN
      END


      SUBROUTINE OPTIM (F0,FIN,T.NTIME,DTIME,F15MIN,HWW,C1,C2,C3)

      INTEGER I, J, K, N, NIT, LU
      REAL  F0, FIN, T, NTIME, DTIME, F15MIN, HWW, C1, C2, C3
      REAL  F(1560), FOLD(12), FNEW(8), OPT(8,11), Y(11)

      COMMON /RTDATA/ F

      DATA N /4/, FOLD /2.54,2.73,2.95,2.97,8*0.0/
```

```
      IF (T.EQ.0.0) GOTO 210
      IF (T.LT.NTIME) GOTO 1000

      J=N-1
      DO 200 I=1,J
200   FOLD(I)=FOLD(I+1)
      FOLD(N)=F15MIN*718.176E-6
210   CONTINUE

C  Use this section of code for the "perfect" flow predictor
C  and comment out the call to subroutine FLOWPR
C       I = INT(T)
C       DO 220 J=1,4
C       FNEW(J)=0.0
C       DO 220 K=1,15
C220    FNEW(J)=FNEW(J)+F(I+15*(J-1)+K)/15.0

      CALL FLOWPR (T,FOLD,FNEW)

      DO 400 I=1,4
      DO 400 J=1,5
      OPT(I,J)=FNEW(I)
      IF (I.EQ.J) OPT(I,J)=OPT(I,J)+1.0
C      IF (J.EQ.5) OPT(I,J)=0.0
400   CONTINUE

      CALL EVAL (HWW,C1,C2,C3,FNEW,OPT(1,1),Y(1),N)
      CALL EVAL (HWW,C1,C2,C3,FNEW,OPT(1,2),Y(2),N)
      CALL EVAL (HWW,C1,C2,C3,FNEW,OPT(1,3),Y(3),N)
      CALL EVAL (HWW,C1,C2,C3,FNEW,OPT(1,4),Y(4),N)
      CALL EVAL (HWW,C1,C2,C3,FNEW,OPT(1,5),Y(5),N)

C         DO 270 LU=9,9,3
C            WRITE (LU,250) (FOLD(I),I=1,N)
C250         FORMAT ('1FOLD = ',8F9.4)
C            WRITE (LU,260) (FNEW(I),I=1,N)
C260         FORMAT (' FNEW = ',8F9.4)
C270      CONTINUE

      NITS=0
      DO 300 I=1,100
      CALL SIMPLX (N,8,OPT,Y,NIT,HWW,C1,C2,C3,FNEW)
C         DO 320 LU=9,9,3
C            WRITE (LU,310) (OPT(J,1),J=1,N),Y(1)
C310         FORMAT (' ',4F10.5,F15.5)
C320      CONTINUE
300   CONTINUE

      F0=OPT(1,1)*92.82776
      NTIME=NTIME+DTIME
      F15MIN=0.0

1000  CONTINUE
```

```
      RETURN
      END


      SUBROUTINE AVGFLW (TIME,FAVG)

      INTEGER I
      REAL A0, A(6), B(6), TIME, FAVG, PI, TEMP

      DATA PI /3.14159/
      DATA A0,A /2.540603, 0.132534, 0.301387, 0.220798,
     1           -0.204561, 0.039848, 0.103949/
      DATA B /-0.464522, -0.546182, 0.079117, 0.110810,
     1        -0.127716, 0.060457/

      FAVG = A0

      DO 100 I=1,6
      TEMP = 2.0*I*PI*TIME/1440.0
      FAVG = FAVG + A(I)*COS(TEMP) + B(I)*SIN(TEMP)
100   CONTINUE

      RETURN
      END


      SUBROUTINE FLOWPR (TIME,FOLD,FNEW)

      INTEGER I
      REAL FOLD(12), FNEW(8), TIME, FBAR(12)

      DO 100 I=1,12
      T = TIME + 15.0*(I-4)
      IF (T.LT.0.0) T=T+24.0
      CALL AVGFLW (T,FBAR(I))
100   CONTINUE

      DO 200 I=1,8
      CALL PR15 (FOLD(I),FBAR(I),FNEW(I))
      FOLD(I+4)=FNEW(I)
200   CONTINUE

      RETURN
      END


      SUBROUTINE PR15 (FOLD,FBAR,FNEW)

      INTEGER I
      REAL FOLD(4),FBAR(5),FNEW,C(4),FC
      DATA C /0.826906, 0.194541, 0.053819, -0.09641/
```

```
      FC = 0.0
      DO 100 I=1,4
100   FC = FC + C(5-I)*(FOLD(I)-FBAR(I))

      FNEW = FBAR(5) + FC

      RETURN
      END


      SUBROUTINE EVAL (HWW,C1,C2,C3,FNEW,FOUT,ERROR,N)

      INTEGER  N
      REAL     HWW, C1, C2, C3, ERROR
      REAL     FNEW(8), HOUT(8), FOUT(8)
      DATA AREA /508.0/

      ERROR = 0.0

      DO 100 I=1,N
      XE=C1+C2*FOUT(I)
100   ERROR=ERROR+ABS(FOUT(I)*XE)

      DO 110 I=1,N
C110   IF (FOUT(I).LT.0.0) ERROR=ERROR+100.0
110   IF (FOUT(I).LT.0.0) ERROR=ERROR+1000.0

      H=HWW
      DO 200 I=1,N
      H = H+1392.4*(FNEW(I)-FOUT(I))/AREA
      HOUT(I)=H
C      IF (H.LT.5.0) ERROR=ERROR+100.0
      IF (H.LT.5.0) ERROR=ERROR+1000.0 + 100.0*(5.0-H)
      IF (H.GT.10.0) ERROR=ERROR+1.0*(H-10.0)
      IF (H.GT.12.0) ERROR=ERROR+100.0+100.0*(H-12.0)
200   CONTINUE

      ERROR=-ERROR

      RETURN
      END


      SUBROUTINE SIMPLX (N,ND,X,Y,NIT,HWW,C1,C2,C3,FNEW)
CCCCCCCCCCCCCCCCCCCC
C
C  Subroutine Name:  SIMPLX - Simplex Optimization
C
CCCCCCCCCCCCCCCCCCCC
C
C  Robert D. Hill
C  Houston, Texas  June 16-18, 1982
C
```

```
CCCCCCCCCCCCCCCCCCCC
C
C  PURPOSE:  This program performs a SIMPLEX optimization.
C            The program attempts to maximize the response.
C            It returns control to the calling program after
C            each iteration.
C            Expansions and contractions are considered as part
C            of each iteration.
C            The calling program determines end conditions.
C
C  ADAPTED FROM:  SIMPLEX OPTIMIZATION IN RESEARCH AND DEVELOPMENT
C                 Department of Chemistry
C                 University of Houston
C                 Houston, Texas  77004  U.S.A
C                 Lecturer:  Dr. Stan Deming
C                 May 14-15, 1979
C
C  LIMITATIONS:  Presently limited to 25 factors.
C
C  VARIABLES:  N   - Number of factors.
C              ND  - Row dimension of "X" in calling program.
C              X   - Factor space.  Each column corresponds
C                      to one evaluation (vertex).
C                      Dimension to (N,N+3) or greater.
C              Y   - Response from each vertex.
C              NIT - Number of iterations.
C                      Set to zero when subroutine first entered.
C              FNC - Name of function which evaluates response.
C
C  SUBROUTINE CALLS:  SORT - Sort responses in decreasing order.
C                     FNC  - Function name which evaluates responses.
C
C  ABREVIATIONS:  P  - Centroid of remaining hyperface.
C                 W  - Worst vertex.
C                 N  - Next to worst vertex (always rejected next).
C                        Not the same as the variable "N".
C                 R  - Reflection vertex.
C                 B  - Best vertex.
C                 E  - Expansion vertex.
C                 CR - Contraction vertex in reflection direction.
C                 CW - Contraction vertex in worst direction.
C
CCCCCCCCCCCCCCCCCCCC
C
C

      INTEGER  N, ND, NIT, I, J
      REAL     X(ND,1), Y(28), P(25), PW(25)
      REAL     FNEW(8), HWW, C1, C2, C3

C
C ----- if first iteration, sort initial data
C
```

```
      IF (NIT.EQ.0) CALL SORT (N+1,ND,X,Y)

C
C ----- Increment iteration counter
C
      NIT = NIT + 1

C
C ----- Calculate P and PW vectors
C
      DO 120 I=1,N
      P(I) = 0.0
      DO 110 J=1,N
110   P(I) = P(I) + X(I,J)
120   P(I) = P(I)/N

      DO 130 I=1,N
130   PW(I) = P(I) - X(I,N+1)

C
C ----- Calculate reflection (R) vector
C
      DO 140 I=1,N
140   X(I,N+2) = P(I) + PW(I)

C
C ----- Evaluate Response at R
C
      CALL EVAL (HWW,C1,C2,C3,FNEW,X(1,N+2),Y(N+2),N)

C
C -----  Decide whether to expand, contract, or keep reflection
C
      IF (Y(N+2).GT.Y(1)) GOTO 300
      IF (Y(N+2).LT.Y(N)) GOTO 400

C
C ----- Keep R as new vertex
C
200   DO 210 I=1,N
      X(I,N+1) = X(I,N)
210   X(I,N)   = X(I,N+2)

      Y(N+1) = Y(N)
      Y(N)   = Y(N+2)

      CALL SORT (N,ND,X,Y)
      RETURN

C
C ----- Try an expansion.
C ----- Calculate expansion vector
C
```

```
300   DO 310 I=1,N
310   X(I,N+3) = X(I,N+2) + PW(I)

C
C ----- Evaluate response at E
C
      CALL EVAL (HWW,C1,C2,C3,FNEW,X(1,N+3),Y(N+3),N)

C
C ----- If E<B, use R
C
      IF (Y(N+3).LT.Y(1)) GOTO 200

C
C ----- Otherwise, keep E
C
320   DO 330 I=1,N
      X(I,N+1) = X(I,N)
330   X(I,N)   = X(I,N+3)

      Y(N+1) = Y(N)
      Y(N)   = Y(N+3)

      CALL SORT (N,ND,X,Y)
      RETURN

C
C ----- Use a contraction.
C ----- If R<W, use CW
C
400   IF (Y(N+2).LT.Y(N+1)) GOTO 450

C
C ----- Otherwise, use CR
C
      DO 410 I=1,N
410   X(I,N+3) = P(I) + 0.5*PW(I)

      CALL EVAL (HWW,C1,C2,C3,FNEW,X(1,N+3),Y(N+3),N)
      GOTO 320

C
C ----- Use CW
C
450   DO 460 I=1,N
460   X(I,N+3) = P(I) - 0.5*PW(I)

      CALL EVAL (HWW,C1,C2,C3,FNEW,X(1,N+3),Y(N+3),N)
      GOTO 320

      END
```

```
      SUBROUTINE SORT (N,ND,X,Y)
CCCCCCCCCCCCCCCCCCCC
C
C  Subroutine Name:  SORT
C
CCCCCCCCCCCCCCCCCCCC
C
C  Robert D. Hill
C  June 16, 1982
C
CCCCCCCCCCCCCCCCCCCC
C
C  PURPOSE:  Internal subroutine called only by "SIMPLX".
C            Sort "Y" and columns of "X" in descending order.
C            Simple "bubble" sort.
C
C  VARIABLES:  N  - Number of factors.
C              ND - Row dimension of "X" in calling program.
C              X  - Factor space, Each column is one vertex.
C              Y  - Response vector.
C
CCCCCCCCCCCCCCCCCCCC
C
C

      INTEGER  N, ND, I, J, K
      REAL     X(ND,1), Y(28), T

      DO 300 I=1,N-1
      DO 200 J=1,N-1
      IF (Y(J).GE.Y(J+1)) GOTO 200
      T = Y(J)
      Y(J) = Y(J+1)
      Y(J+1) = T
      DO 100 K=1,N
      T = X(K,J)
      X(K,J) = X(K,J+1)
      X(K,J+1) = T
100   CONTINUE
200   CONTINUE
300   CONTINUE

      RETURN
      END
```

```
*  PROGRAM REACTOR
*
*  ROBERT D. HILL
*  DECEMBER 23, 1983
*
*  THIS PROGRAM CALCULATES SUSPENDED SOLIDS CONCENTRATIONS
*  IN THE REACTOR CONFIGURATION OF THE SAGEMONT WWTF FOR
*  ARBITRARY FLOW INPUTS.
*
FIXED HR,MNT
*
INITIAL
*
CONSTANT  V1=785.0,  V2=785.0,  V3=1032.0,  V4=688.0
CONSTANT  V5=688.0,  V6=382.0,  V7=382.0,  V8=101.0
CONSTANT  V9=101.0,  V10=101.0,  V11=101.0,  V12=101.0
*
CONSTANT  VR1=101.0,  VR2=101.0,  VR3=101.0
CONSTANT  VR4=101.0,  VR5=101.0,  VR6=191.0
CONSTANT  VR7=191.0
*
INCON  IX1=6867.0,  IX2=6674.0,  IX3=6620.0,  IX4=6630.0
INCON  IX5=6675.0,  IX6=6709.0,  IX7=6752.0,  IX8=6764.0
INCON  IX9=6776.0,  IX10=6790.0,  IX11=6803.0,  IX12=6818.0
*
INCON  IXR1=11161.0,  IXR2=11088.0,  IXR3=11085.0
INCON  IXR4=11090.0,  IXR5=11079.0,  IXR6=11016.0
INCON  IXR7=10925.0
*
*
DYNAMIC
*
PROCEDURE F0,FR,FF,FW,X0,XR,MXW,MX10=INPUT(TIME,KEEP)
      IF (KEEP.NE.1) GOTO 200
      IF (TIME.EQ.0) GOTO 110
100   IF (TIME.LT.NTIME) GOTO 200
      F0=AF0
      FR=AFR
      FF=AFF
      FW=AFW
      X0=0.0
      XR=AXR
      MXW=AX2
      MX10=AX4
110   READ (5,120) HR,MNT,AX4,AXR,AXE,AFE,AFR,AX2,AFW,AFF
120   FORMAT (I4,I3,F7.1,F8.1,2F7.3,F6.3,F7.1,F6.1,F6.3)
      NTIME=HR+MNT/60.0
*  CONVERT MGD TO CU M/HR
      AFE=AFE*157.71
      AFR=AFR*157.71
      AFF=AFF*157.71
```

```
*   CONVERT GPM TO CU M/HR
        AFW=AFW*0.2271
        IF (AFW.LT.1.0) AFW=0.0
*   CALCULATE INFLUENT FLOW RATE FROM EFFLUENT AND WASTE
        AF0=AFE+AFW
        GOTO 100
200     CONTINUE
ENDPROCEDURE
*
*   CALCULATE FLOWS
        F1=F0+FR-FF
        F2=F1
        F3=F2
        F4=F3-FW
        F5=F4
        F6=F5+FF
        F7=F6
        F8=F7
        F9=F8
        F10=F9
        F11=F10
        F12=F11
*
*   MASS BALANCES
        DX1=((F0-FF)*X0+FR*XR7-F1*X1)/V1
        DX2=(F1*X1-F2*X2)/V2
        DX3=(F2*X2-F3*X3)/V3
        DX4=(F3*X3-(F4+FW)*X4)/V4
        DX5=(F4*X4-F5*X5)/V5
        DX6=(F5*X5+FF*X0-F6*X6)/V6
        DX7=(F6*X6-F7*X7)/V7
        DX8=(F7*X7-F8*X8)/V8
        DX9=(F8*X8-F9*X9)/V9
        DX10=(F9*X9-F10*X10)/V10
        DX11=(F10*X10-F11*X11)/V11
        DX12=(F11*X11-F12*X12)/V12
*
        DXR1=FR*(XR-XR1)/VR1
        DXR2=FR*(XR1-XR2)/VR2
        DXR3=FR*(XR2-XR3)/VR3
        DXR4=FR*(XR3-XR4)/VR4
        DXR5=FR*(XR4-XR5)/VR5
        DXR6=FR*(XR5-XR6)/VR6
        DXR7=FR*(XR6-XR7)/VR7
*
*   INTEGRAL CALLS
        X1=INTGRL(IX1,DX1)
        X2=INTGRL(IX2,DX2)
        X3=INTGRL(IX3,DX3)
        X4=INTGRL(IX4,DX4)
        X5=INTGRL(IX5,DX5)
        X6=INTGRL(IX6,DX6)
        X7=INTGRL(IX7,DX7)
```

```
      X8=INTGRL(IX8,DX8)
      X9=INTGRL(IX9,DX9)
      X10=INTGRL(IX10,DX10)
      X11=INTGRL(IX11,DX11)
      X12=INTGRL(IX12,DX12)
*
      XR1=INTGRL(IXR1,DXR1)
      XR2=INTGRL(IXR2,DXR2)
      XR3=INTGRL(IXR3,DXR3)
      XR4=INTGRL(IXR4,DXR4)
      XR5=INTGRL(IXR5,DXR5)
      XR6=INTGRL(IXR6,DXR6)
      XR7=INTGRL(IXR7,DXR7)
*
      MLSS2=(X4+X5)/2.0
*
*
TERMINAL
TIMER FINTIM=48.0, OUTDEL=0.10, PRDEL=1.0, DELMAX=0.008333333
PRTPLT F0,FR,FF,FW,X0,XR,MXM,MX10
OUTPUT X12(3000.0,8000.0),MX10(3000.0,8000.0)
OUTPUT MLSS2(3000.0,8000.0),MXW(3000.0,8000.0)
OUTPUT X4(3000.0,8000.0),MXW(3000.0,8000.0)
OUTPUT X5(3000.0,8000.0),MXW(3000.0,8000.0)
PRINT X1,X2,X3,X4,X5,X6,X7,X8,X9,X10,X11,X12,...
XR1,XR2,XR3,XR4,XR5,XR6,XR7
END
INPUT
   0  1 6446.5 10954.7  0.000  2.450 3.673 6385.9   0.6 0.000
   0  2 6520.3 10928.1  0.000  2.471 3.698 6350.8   0.6 0.000
   0  3 6500.4 10946.9  0.000  2.470 3.641 6354.3   0.6 0.000
   0  4 6496.5 10787.5  0.000  2.454 3.737 6396.9   0.6 0.000
   0  5 6557.8 10556.3  0.000  2.387 3.682 6376.1   0.6 0.000
   0  6 6467.6 10460.9  0.000  2.293 3.675 6370.3   0.6 0.000
   0  7 6490.6 10693.0  0.000  2.282 3.622 6507.8   0.6 0.000
   0  8 6520.3 10939.8  0.000  2.290 3.636 6365.6   0.6 0.000
   0  9 6508.6 10941.4  0.000  2.352 3.643 6374.6   0.6 0.000
   0 10 6536.3 10954.7  0.000  2.362 3.579 6374.2   0.6 0.000
                                   .
                                   .
                                   .
          2880 DATA POINTS (48 HOURS OF ONE MINUTE DATA)
                                   .
                                   .
                                   .
  47 59 6303.1 11476.6  3.213  1.599 2.690 6345.7   0.7 0.000
  48 00 6269.9 11394.5  3.108  1.599 2.656 6375.4   0.6 0.000
ENDINPUT
STOP
```

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