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Evaluation of Alternative Unstructured Kinetic Models for Simulating the Dynamics of the Activated Sludge Process

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Abstract. Four unstructured kinetic models were used for simulating the transient response of the conventional activated sludge to shock loadings. The aeration tank was assumed to be a continuous stirred tank reactor while the settler dynamics were simulated as a simple first order lag. Parameter identification was exercised in predicting the models substrate and biomass concentrations as compared to published experimental results. The inhibitory endogenous metabolism model gave relatively good qualitative results. The inadequacy of these simple models in predicting some important aspects of the unsteady state operation is discussed.

1. Introduction

Both practical experience and in-depth analysis have shown that the biological wastewater treatment plants seldom, if ever, operate at steady state. The ASP's usually operate under transient conditions because industrial plant operations [1], which constitute the source of wastewater inflows to ASP's, consists of batch operations, store-and-dump waste procedures as well as occasional spills, all of which contribute to periodic variations in the strength and flow of wastewater streams. Also, municipal flows can vary diurnally from 50 to 150 percent of the average flow. In addition household wastewaters are usually discharged in sewers at approximately the same daily periods. Gaber and Elnashaie [2] investigated theoretically the possibility of conveyance/sewage treatment, however their work is confined to steady state analysis and steady state response. The physico-chemical characteristics of the process has been well analyzed recently by Torruelles [3]. Holmberg [4] built a simple dynamic model of the activated sludge process for use with microprocessor-based online estimation and control of the system to compensate for fluctuations in the feed due to large daily variations in organic and hydraulic loading rates [5]. Such disturbances constitute shock loadings to the ASP which usually translate into a lower efficiency of the treatment [6].

As an indication of the extent to which this phenomena has been recognized, a new approach to the problem of control of ASP has been reported in the literature [7] which assumes that the process is in fact a periodic process. That is, the incoming periodic pattern of feed fluctuations is explicitly recognized when designing an operation strategy for the unit.

Engineers entrusted with the design of biological treatment facilities have, in the past, ignored transient loadings and have tried to improve the performance of ASP's by using large equalization basins and overdesigned treatment plants [5].

Although, an equalization basin preceding the ASP would provide some dampening of the concentration peaks, this process arrangement does not exist for a large number of operating plants and does not always solve the problem adequately. Overdesigned treatment plants may also provide a way of minimizing the impacts of feed perturbations but again this does not provide a satisfactory solution on an economical as well as an operational basis [6].

The fundamental problem then lies in the determination of automatic process control strategies which can adequately overcome the undesirable effect of such disturbances. One of the most efficient techniques for developing such control strategies is through the use of dynamic modeling and computer simulation.

Many such models exist in the literature but unfortunately most have been used for design purposes or for predicting steady state behaviour.

Studies on dynamic simulation [6] have suffered in general from one or more of the following:

(a) They often bear little relationship to real plant operation.
(b) They are rarely or poorly validated by experimental data.
(c) They tend to be highly structured thus necessitating the determination (experimentally or empirically) of many parameters. Thus, large amounts of data are needed to provide a comprehensive and accurate model.
(d) These high-power models require considerable computation times on large main frame computers and in some instances hybrid computer systems [8].

In this study we simulate the typical ASP by four low-order growth kinetic models and then test the validity of each of these models against published experimental data.

Past work [1] has shown that such simplifications introduced by a reduced-order model are largely acceptable. Furthermore, the choice of a low-order model facilitate the implimentation of the resulting algorithms on small computer structures such as microprocessors.

2. Models Development

Fig. 1. shows the flow diagram of a typical activated sludge plant. Such a process is one of aerobic digestion in which excess air, and hence oxygen, is introduced in the aeration tank so that there is never a deficiency of oxygen available for the aerobes [9]. For this reason we have assumed that in developing our models that the microbial growth is controlled by organic substrate limitation rather than oxygen limitation. In the ASP much of the soluble and colloidal organic material remaining after primary sedimentation are metabolized by a diverse group of micro-organisms, mainly bacteria, which is responsible for oxidizing most of the organic material to extra-cellular mass, CO_2 and H_2O.

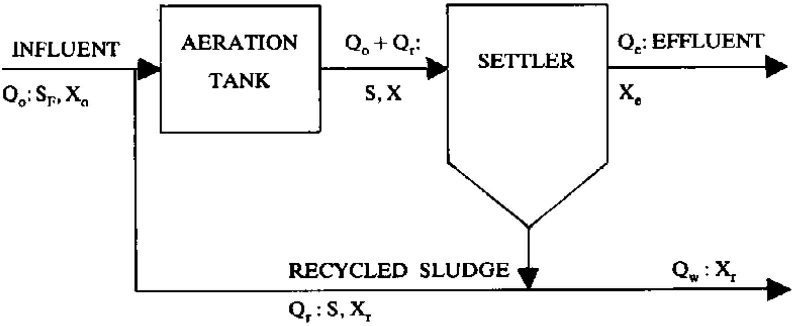


Fig. 1. The conventional activated sludge process

Thus, the biological reaction in the aeration tank is autocatalytic and is the basis of the growth kinetics in the four models considered in this study:

$$\text{Organic pollutants} + O_2 \xrightarrow{\substack{\text{Aerobic}\\ \text{organisms}}} \text{Aerobic organisms} + CO_2 + H_2O \qquad (1)$$

The activated sludge tank is followed by a sedimentation tank (also referred to as a thickener or settler) in which the mixed liquor suspended solids (MLSS) is settled out as sludge and recycled to the aeration tank. This recycling is necessary because ASP's are characterized by optimum MLSS quantities which should be maintained if easily settled sludges are to be obtained. Such levels of MLSS can only be achieved by recycling [9].

In developing our models we have decoupled the primary settler (not shown in the flow diagram), which precedes the aeration tank, from the secondary portion of the model. However, there is a high level of interaction between the aeration basin and the solids-liquid separator [8]. This is so because the performance of the solids-liquid separator will determine to a large extent the concentration of the MLSS that can be maintained in the aeration basin, as well as the quantity of suspended solids in the process effluent [8].

3. Assumptions

1. The aeration tank is treated as a continuous stirred tank reactor (CSTR).
2. The process is assumed homogeneous.
3. Heat effects due to reaction or cooling due to evaporation are neglected.
4. Physical properties of the fluids such as viscosity and density are constant.
5. Microbial growth in the thickener is neglected.
6. No distinction is made between soluble and suspended substrate [8].
7. The recycle substrate concentration is equal to the substrate concentration in the activated sludge reactor effluents(S). In fact, it has been indicated in the literature [10] that taking S_R as an independent variable or as equal to S or as negligible lead to slight differences in the solution of the balance equations.
8. The biomass recycle concentration (x_R) is directly proportional to reactor effluent biomass concentration (x) at steady state operation. When the system is disturbed, x_R lags x by a first order time lag τ_s. This τ_s incorporates general overall time lags between reactor outlet point and recycle stream introduction point to reactor, thus both thickener as well as pipeline transportation lags are included.
9. The biokinetics are controlled by organic substrate, rather than oxygen, limitation.

4. Substrate Removal Rate Kinetic Models

Four micro-organisms growth rates have been considered and these are:

4.1. The Monod Model

The growth rate function (μ) is expressed as function of substrate concentration as follows:

$$\mu = \frac{\mu_m \ S}{K_s + S} \tag{2}$$

where μ = specific growth rate (t^{-1}), S = substrate concentration
(mg/lit), μ_m = maximum specific growth rate (t^{-1}) and
K_s = saturation constant (mg/lit).

It is evident from the Monod relationship that the specific growth rate follows first order kinetics when S is very low and follows zero order kinetics when S is very high.

For non-inhibitory substrates several citations in the literature have indicated that the Monod equation adequately expressed the growth rate-substrate relationship [5,10,11,12,13] in terms of biomass concentration:

$$\frac{dx}{dt} = \mu x \tag{3}$$

where x is biomass concentration in mg/lit

4.2. Monod-Endogenous Metabolism

Endogenous metabolism means that there are reactions in cells which consume cell substance. This is a modification to the Monod relationship to account for the fact that the observed evidence of marked decline in cell concentration at low dilution rates in continuous cultures of micro-organisms [9]. Thus we might write:

$$\frac{dx}{dt} = \mu x - k_d x \tag{4}$$

The additional term ($k_d x$) has sometimes also been interpreted as a cell death or decay rate constant, and k_d has the same as μ, (t^{-1}).

4.3. Inhibition Model (Haldane Equation)

For certain cultures of micro-organisms (*e.g.* phenol wastes [10]) it has been observed that the actual reaction rate - substrate concentration curve does not increase to μ_m and then stays constant, but in fact it increases, then remains constant for a certain time, then decreases with the increase in substrate concentration. This represents inhibition of the biological reaction by the substrate. In such cases:

$$\frac{dx}{dt} = \mu x$$

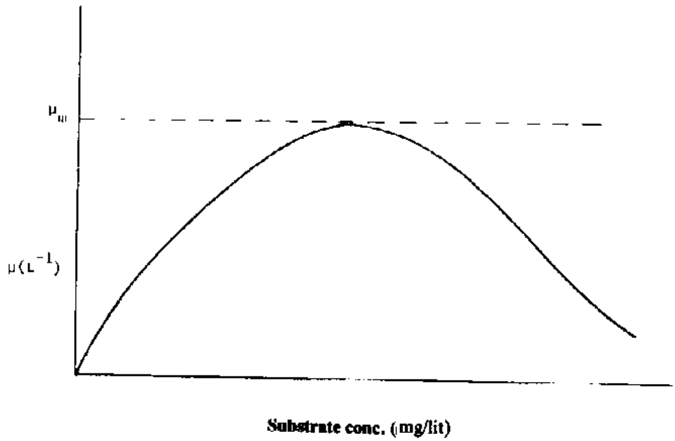
where

$$\mu = \frac{\mu_m S}{K_s + S + \frac{S_2}{K_2}} \tag{5}$$

4.4. Inhibition and Endogenous Metabolism

This model has also been referred to in the literature as the inhibition and decay model [9]. For this type of model the rate of increase of biomass concentration is given by the relationship:

$$\frac{dx}{dt} = \mu x - k_d x = \frac{\mu_m S}{K_s + S + \frac{S_2}{K_2}} x - k_d x \tag{6}$$



Inhibition kinetics

Regardless of whether inhibition is caused by excess substrate or products o reaction concentration, the net effect is a reduction in viable organisms - that is, those organisms which are effective in causing substrate consumption [9].

5. Material Balance

With reference to Fig. 1, the following mass balance applies to any component in the aeration tank:

Accumulation = Rate of production + Rate of addition − Rate of removal

Let:

x = biomass concentration in aeration tank, mg/lit.

x_0 = biomass concentration in feed, mg/lit.

x_r = biomass concentration in recycle stream, mg/lit.

Q_0 = F = volumetric feed flow rate, lit/hr.

Q_r = Recycle flow rate, lit/hr.

V = aeration tank volume, lit.

S = substrate concentration in aeration tank, mg BOD/lit.

$\tau = \frac{V}{F} =$ aerator residence time, hr.

Y = yield coefficient = $\frac{\text{mass of micro-organisms produced}}{\text{mass of substrate utilized}}$

R = Recycle ratio = $\frac{Q_r}{Q_0}$

α = $\frac{x_r}{x}$

A biomass balance for the Monod kinetics model gives:

$$V\frac{dx}{dt} = Q_0 x_0 + Q_r x_r - (Q_0 + Q_r)\, x + V\mu x \tag{7}$$

Dividing by V and assuming $x_0 = 0$ and noting that $\tau = \frac{Q_0}{V}$:

$$\frac{dx}{dt} = \frac{Q_r}{V} x_r - (\tau + \frac{Q_r}{V})\, x + \mu x \tag{8}$$

$$\text{Let recycle ratio} = \frac{Q_r}{Q_0} = R \tag{9}$$

$$\frac{x_r}{x} = \alpha \tag{10}$$

The coefficient α can be assumed constant for steady state operation or a time-dependent function to account for the time lag due to pipeline and settler dynamics. Assuming that the transportation lags together with settler dynamics can be modeled as a first order lag then:

$$\frac{x_r}{x} = \alpha = 4\,x(t_0) + 4(x - x(t_0))\ (1 - e^{\frac{t}{\tau_s}}) \tag{11}$$

where $x(t_0)$ is value of x at t = and τ_s is a time constant accounting for settler residence time and transportation lags in the pipeline and t is time elapsed after onset of disturbance.

The substrate concentration at any time can be expressed as follows:

$$V\frac{dS}{dt} = Q_0\ S_F + RQ_0S - (1 + R)\,Q_0S - V\frac{\mu x}{Y} \tag{12}$$

$$\therefore \frac{dS}{dt} = \tau\,(S_F - S) - \frac{\mu x}{Y} \tag{13}$$

The substrate removal efficiency (η) is an indicator of process performance and is expressed as follows:

$$\eta = \frac{S_F - S}{S_F} \tag{14}$$

It is worth noting that the expression for settler dynamics is probably an oversimplification whose validity should be tested against experimental measurements. Highly structured and elaborate settler models have been presented in the literature [5,10].

Most of the dynamic thickener models proposed in the literature [14] have a common foundation of numerous models in the solids flux theory of Dick. [15]. In these models the basic concept is that the total flux at any location inside a thickener is the sum of the batch settling flux and the withdrawal flux induced by sludge withdrawal from the bottom. Such dynamic models are developed by dividing the thickener into several completey mixed layers or zones. A dynamic model of such a structure would give due consideration to the three main functions of the solids-liquid separator, namely those of thickening, clarification and sludge storage [8].

Table 1 summarizes the four mathematical models as used in this study.

Table 1. The Mathematical models and their characterizing equations

	dx/dt	ds/dt	μ
Monod	$(\theta + \mu)X$	$\tau(S_F - S) - \mu\frac{x}{y}$	$\frac{\mu_m \quad S}{K_s \quad + S}$
Monod endogenous	$(\theta + \mu - K_d)X$	$\tau(S_F - S) - \mu\frac{x}{y}$	$\frac{\mu_m \quad S}{K_s \quad + S}$
Inhibitory	$(\theta + \mu)X$	$\tau(S_F - S) - \mu\frac{x}{y}$	$\frac{\mu_m \quad S}{S + K_s + S^2/K_2}$
Endogenous inhibitory	$(\theta + \mu - K_d)X$	$\tau(S_F - S) - \mu\frac{x}{y}$	$\frac{\mu_m \quad S}{S + K_s + S^2/K_2}$

where $\theta = \tau(R(\alpha - 1) - 1)$

6. Simulation Results and Discussion

The numerical values of the models parameters at steady state operation were; $\mu_m = 0.55\ h^{-1}$, $K_s = 120.0$ mg BOD/L, $K_2 = 500.0$ mg BOD/L, $Y = 0.55$, $K_d = 0.0025\ h^{-1}$, $S_F = 100$ mg BOD/L, $\alpha = 4$, $R = 0.25$, $\tau = 10$ h and $\tau_s = 4$h.

The four mathematical models were used to simulate the dynamic response of an activated sludge plant to step perturbations in the feed substrate concentration (S_F), the recycle ratio (R) and aeration tank retention time (τ). Three types of wastes were considered in this study, namely domestic waste, chemical wastes and soybean waste. The steady state efficiencies for each of the four models for the different wastewaters are shown in Table 2.

Table 2. The steady state efficiencies for each of the four models for the different wastewaters

Model	Domestic	Chemical	Soybean
Monod	94.286	96.623	82.368
Monod Endogenous	93.684	96.230	77.857
Inhibitory	94.250	96.027	81.520
Endogenous Inhibitory	94.000	95.812	76.850

The system transient response to a step load was plotted as reactor substrate concentration (s), biomass concentration (x) and substrate removal efficiency (η) against time. In the initial stage of the study the settler dynamics were ignored and the transients were simulated assuming that there was an instantaneous correspondence between x_r and x. The thickener dynamics were subsequently incorporated in the overall model as a first order lag.

We don't have space here to show all the results of the transient studies for each of the four models, but they can be found in Mudallal [16]. The four models gave qualitatively similar transient responses. In this paper we present the dynamic behaviour based on the Monod model only.

6.1. The Response to a Step Input to S_F

The model was subjected to four different step size increases in S_F. These step sizes were 35, 50, 65 and 80 mg BOD/lit., for the different types of wastewaters. The transient response to these disturbances showing the variation of S, x and η versus time are plotted in Figs. 2, 3 and 4 for a waste feed consisting of domestic wastewaters.

As is evident from Fig. 2, a step up in S_F causes an immediate sharp increase in S until a peak is reached. Thereafter a gradual continuous decrease in S takes place until a final steady state is reached. The initial short horizontal line indicates steady state operation before imposition of disturbance.

The steady state relationships are obtained by setting $\frac{dx}{dt} = \frac{ds}{dt} = 0$,

$$S = \frac{K_s \tau (R(\alpha - 1) - 1)}{\tau(1 - R(\alpha - 1)) - \mu_m} \tag{15}$$

$$x = \frac{\tau Y(S_F - S)}{\mu} \tag{16}$$

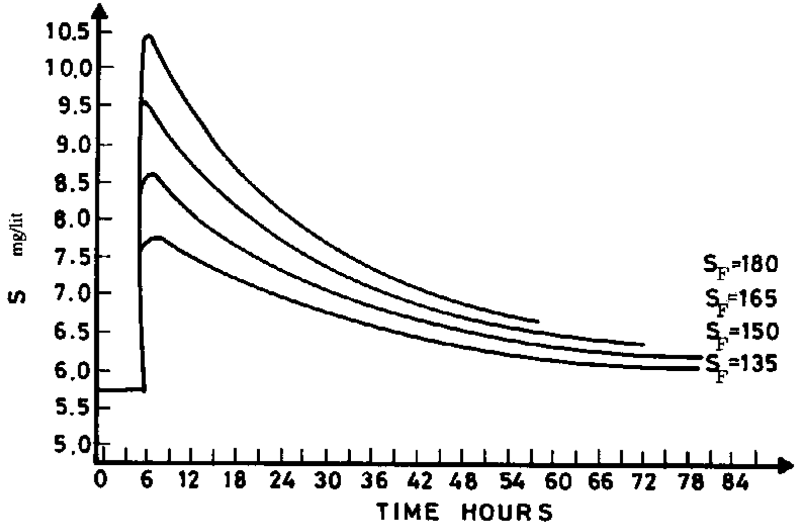


Fig. 2 . Effect of the influent substrate concentration step up on the system substrate concentration

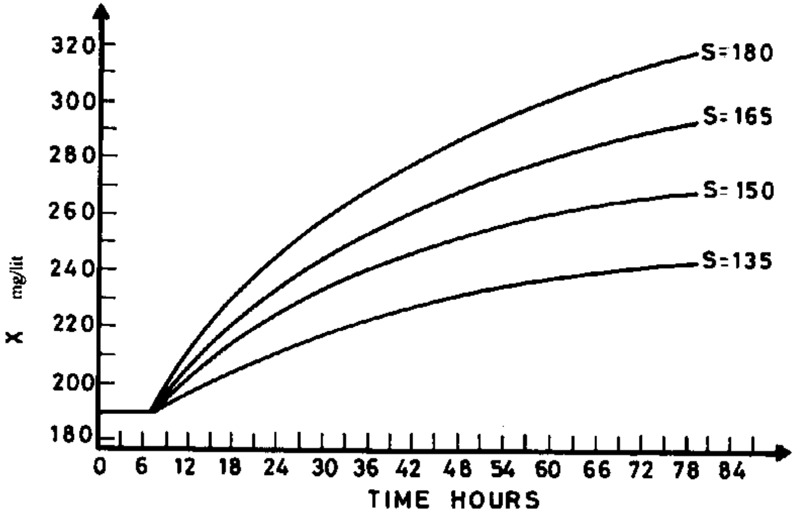


Fig. 3 . Effect of the influent substrate concentration step up on the system biomass concentration

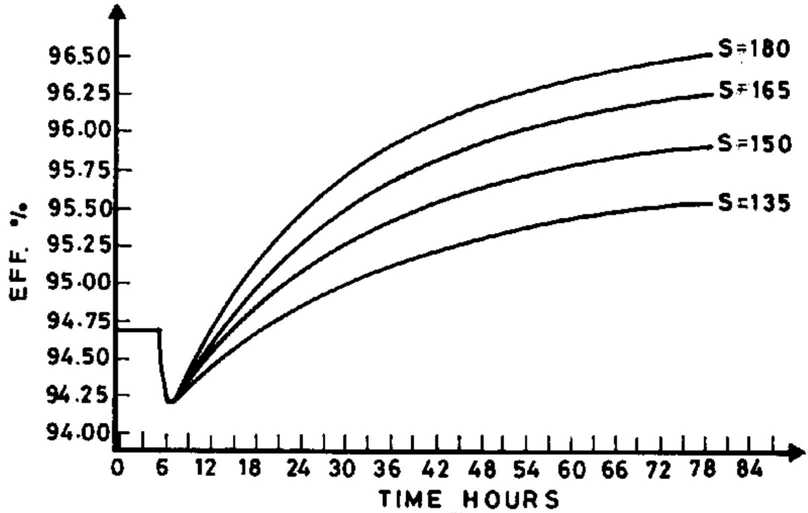


Fig. 4 . Effect of the influent substrate concentration step up on the substrate removal efficiency

However, due to practical limitations, x cannot increase indefinitely, as predicted by equation (16), as S_F is increased. The independence of S w.r.t. S_F as predicted by equation (15) also predict that the efficiency η as given by the relationship

$$\eta = \frac{S_R - S}{S_R}$$

would increase indefinitely as S_F is increased. This prediction can only be made within certain limits as the system efficiency should start to fall when S_F exceeds certain limits.

The occurrence of the sharp initial increase and the overshoot can be explained in terms of the time lag between the sudden increase in the biodegradable wastes concentration (S) and the resultant increase in the biological cell metabolism.

When S is increased, cell growth increases (up to a certain maximum) according to the relationship:

$$\mu = \frac{\mu_m}{K_s + S}$$

and hence an increase in biomass results. As a result of the increase in biomass population, the substrate is biodegraded and hence S starts to fall after reaching a peak. As S decreases, the increase in x slows down until a balance between S and x is reached according to the steady state relationship. The biomass concentration behaviour is as shown in Fig. 3. The biomass does not increase as suddenly as that of S because the biomass takes some time to accomodate to the new change. As predicted by the steady state relationships between S and x an increase in S_F should eventually result in a new final higher steady state value of x.

As shown in Fig. 4 the substrate removal efficiency, due to its inverse dependence on the ratio S/S_F, decreased sharply initially as S_F was stepped up. Thus a step up in S_F results in an immediate deterioration in plant performance. However, once the micro-organisms start multiplication, the resultant biological oxidation of the organic waste resulted in a lower S and hence the efficiency η picked up again until a new steady state is reached. The new value of η is different from its initial value because although the initial and final values of S are identical and S_F has been changed permanently to a new value.

Figures 5, 6 and 7 show the effect of stepping up the recycle ratio (R) on system dynamics. The beneficial effect of increasing R on system performance is indicated by examining Figs. 5 and 6. This suggests the use of R as a control variable (a manipulated variable) in a feed forward control scheme to counteract the transient detrimental effect of increases in S_F. However in practical terms R can be manipulated within a certain range of values; the upper limit of which is set by the pumping, flow, power and aerator size constraints. Thus apart from the technical limitations, the overall higher capital and operating costs will have to be weighed against the resultant increase in substrate removal efficiency. In fact such an optimization exercise might take into consideration building two or more smaller units rather than a single very large and complex one.

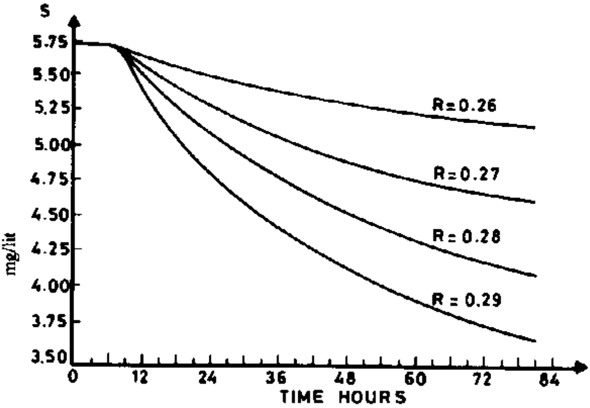


Fig. 5 . Effect of recycle ratio step up on the system substrate concentration

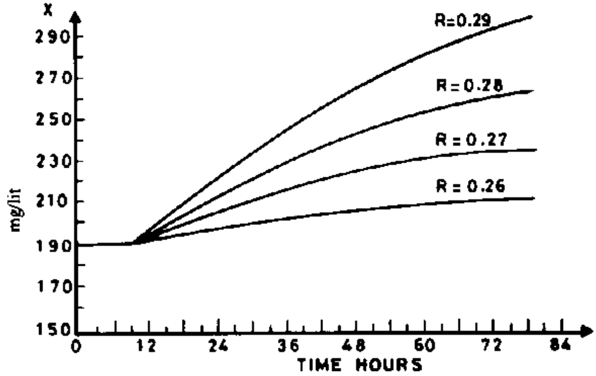


Fig. 6 . Effect of recycle ratio step up on the system biomass concentration

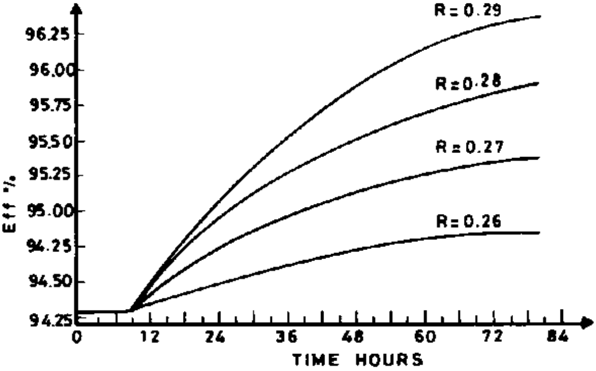


Fig. 7 . Effect of the recycle ratio step up on the system substrate removal Eff

It is worth noting that no peaks are predicted by this transient simulation. This is due to the fact that the additional recycle stream flow rate will add additional microbial cells as well as substrate to the reactor. The substrate addition will be relatively small compared to that of S_F and thus will not have a comparable effect to that of stepping up S_F.

Figures 8, 9 and 10 show the effect of a step down in τ (*i.e.* a step up in Q_0) upon the system performance. S_F was maintained constant. As τ was decreased the system substrate concentration increased and the system efficiency (η) decreased. This is so

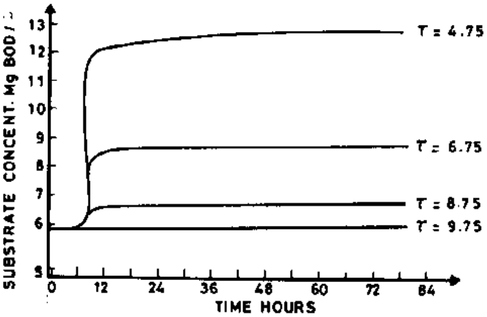


Fig. 8 . Effect of the fresh feed step up on the system substrate conc.

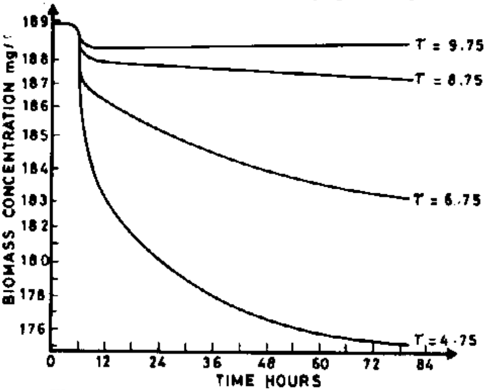


Fig. 9 . Effect of the fresh feed step up on the system biomass concentration

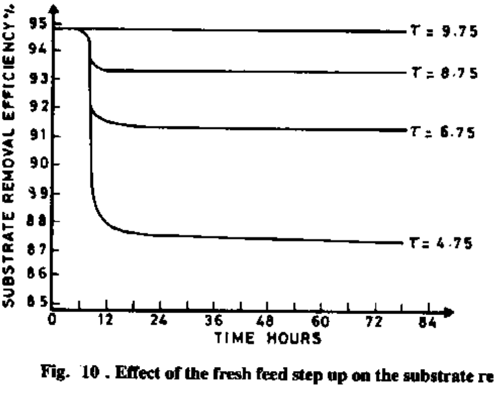


Fig. 10 . Effect of the fresh feed step up on the substrate removal Eff

because when Q_0 is increased the biological mass in the aerator has less time to effect the biodegradation of the substrate. If there is no recycle and Q_0 is stepped up sufficiently, complete cells WASHOUT occurs and the organic waste passes through system without undergoing any biological treatment.

6.2. Effect of Settler Dynamics

The settler dynamics were simulated as a first order lag. At steady state x_r and x are related by the following relationship:

$$x_r = 4\,x$$

when the system is disturbed we simulated the time lag due to settler residence time (as well as transportation lags) by the relationship:

$$x_r = 4\,x\,(t_0) + 4(x - x(t_0))\,(1 - e^{-(t/\tau_s)})$$

where:

x = biomass conc. in reactor effluent at time t.

x_r = recycle biomass concentration at time t.

$x(t_0)$ = biomass conc. in reactor effluent upon onset of distrurbance (*i.e.* at $t = 0$)

τ = residence time (assumed = 4 hours).

t = time elapsed after onset of disturbance.

Figures 11, 12 and 13 show the effect of incorporating settler dynamics upon system transients when S_F is stepped up.

6.3. Comparison of Simulation Results with Published Data

In order to indicate the qualitative validity of the models, comparisons were made between the dynamic simulation results and comparable data published in the literature.

The values of the biokinetic parameters μ_m, K_s, K_2, Y and K_d were obtained from several sources and are considered average values which do not represent any particular activated sludge wastewater treatment plant. Thus using average values for the above five biokinetic constants and the appropriate values for the selectable engineering parameters τ, τ_s, S_F, α and R we simulated the dynamic response of an experimental laboratory size activated sludge unit to transient loading conditions. Fig. 14 shows a drawing of the activated sludge unit used to generate the transient experimental data as reported in technical report No. CRWR 37 [5] published by the Centre for Research in Water Resources (Univ. of Texas). The wastewater was domestic waste and the transient load was a pulse disturbance in the feed organic waste concentration S_F. In making these simulations, several computer runs were carried out to see if changes in the selectable engineering variables (within the allowable range of values reported for the experimental runs) can improve the fit between the simulation predictions and experimental results.

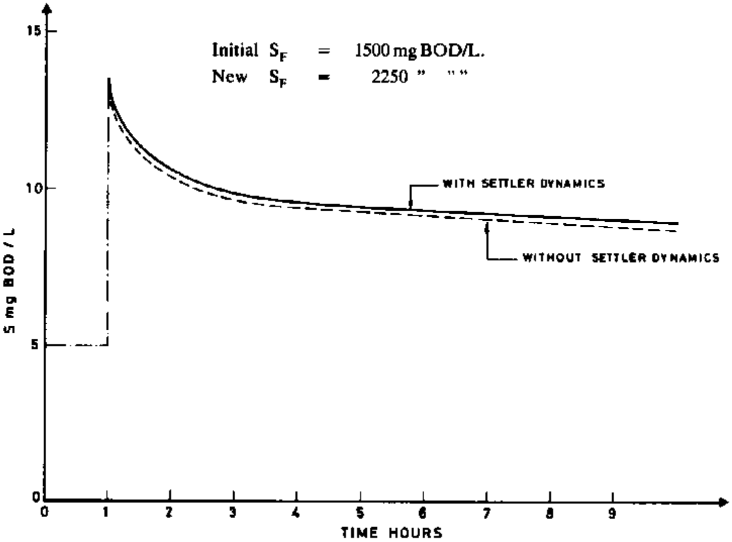


Fig. 11 . Effect of settler dynamics on substrate concentration

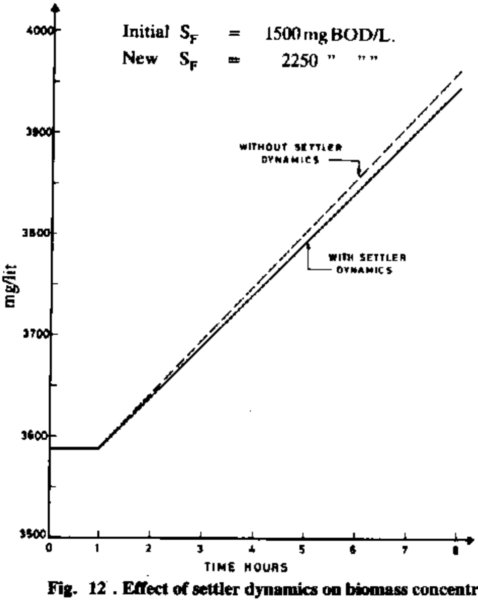


Fig. 12 . Effect of settler dynamics on biomass concentration

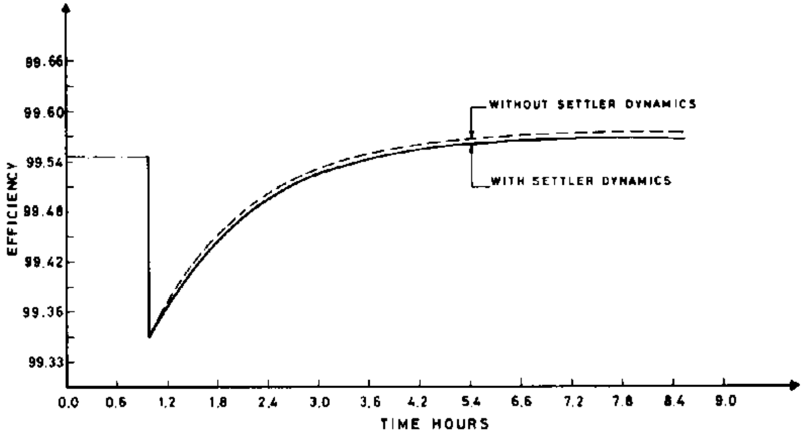


Fig. 13 . Effect of settler dynamics on efficiency

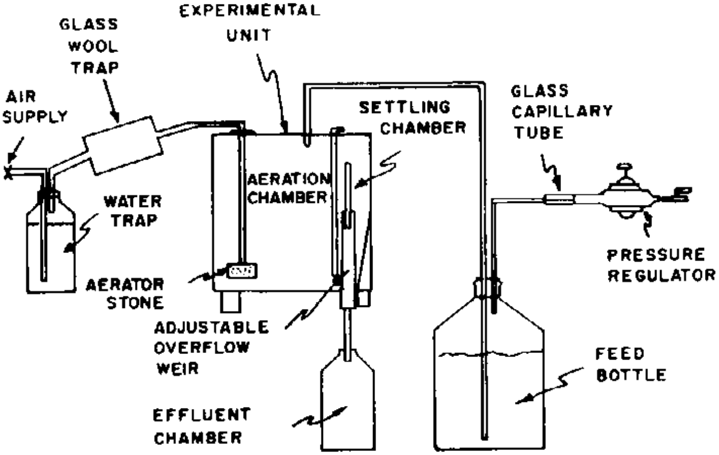


Fig. 14 . Lab-scale activated sludge unit flow diagram

Figure 15 shows the simulation results for the four kinetic models as well as the experimental results. As is seen from these plots the four kinetic models gave qualitatively similar results to each other and to the experimental results. However the Inhibitory + Endogenous metabolism model gave a closer fit to experimental results especially in predicting the final steady state conditions. The experimental results exhibited a transient lag and an overall delay in the transient response.

The time lags can in fact be explained in terms of kinetic as well as engineering considerations: The four kinetic models under consideration are known as unstructured models and have the drawbacks of not accounting for the lag phase (see Fig. 16) and they give us no insight into the variables which influence growth.

Also, they make no attempt to utilize or recognize recent knowledge about cellular metabolism and regulation [12]. Busby and Andrews [8] have noted that lags were observed with some sludges. They presented a structured model which allows for such lags [7]. In so doing they divided the sludge mass into three, rather than one mass. These masses are *stored mass* (which consists of dissolved substrate, suspended solids and colloidable biodegradable organic matter), *active mass* (which is the viable biological mass responsible for the biodegradation of organic matter) and *Inert mass* (which is the result of endogenous metabolism and decay). Thus by making the assumption that all substrate, whether dissolved or suspended, must pass through a

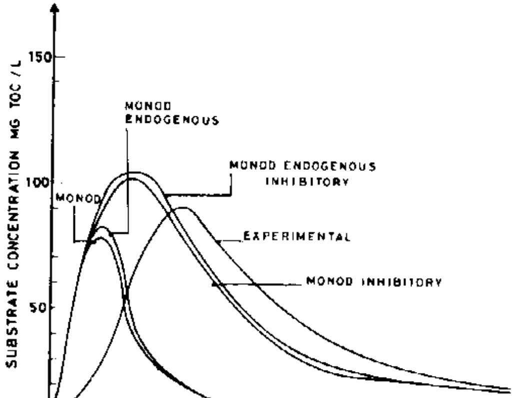


Fig. 15 . 10X Pulse transient response (without thickener dynamics)

$\tau = 6.4$ hours $R = 0.25$

Domestic wastes

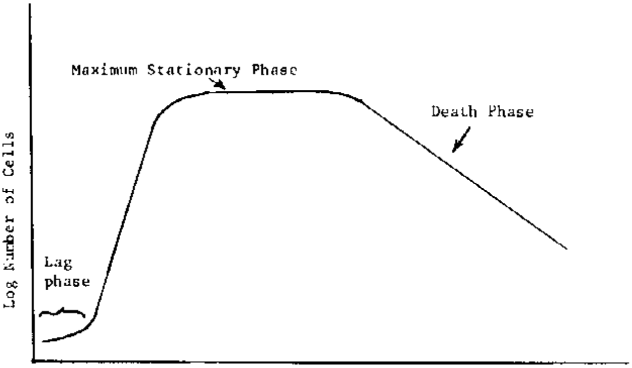


Fig. 16 . Typical batch growth curve of micro-organisms

storage phase as part of the total sludge mass before being used by micro-organisms, they provided a basis for a variable time lag for any transients in substrate concentration which is of considerable importance for predicting the dynamic behaviour of the process. In addition to allowing for transient lags, such structured models take into consideration the fact that a fraction of the total biological mass consists of inert or dead cells. In the simple Monod model all cells are assumed viable and this might explain the discrepancy between the high substrate concentration as given by the experimental results, especially as the system stabilize again after the shock disturbance wears out, and the Monod simulation predictions.

In addition to the deficiency of the four models in accounting for time lags, such time delays can also be attributed to the effect of the dead time due to the transportation lag in transmitting the changes in S_F from the stock solution in the feed bottle to the aeration tank. This is particularly relevant when we note the relatively low flow rate of 30 lit/day used in the experimental studies. Fig. 17 shows that a better fit to experimental data is obtained when a dead time delay of one hour is assumed.

Other plausible explanations for the discrepancy between the simulated and experimental results can (in general) be attributed to:

a) The assumption of zero input biomass concentration ($x_0 = 0$) in the model does not accurately reflect experimental conditions.

b) Incorrect value of the kinetic parameters. As outlined earlier the values of μ_m, K_s, K_2, K_d and Y were taken from the literature whereas the ideal approach is to determine such parameters by laboratory studies on the particular culture in hand.

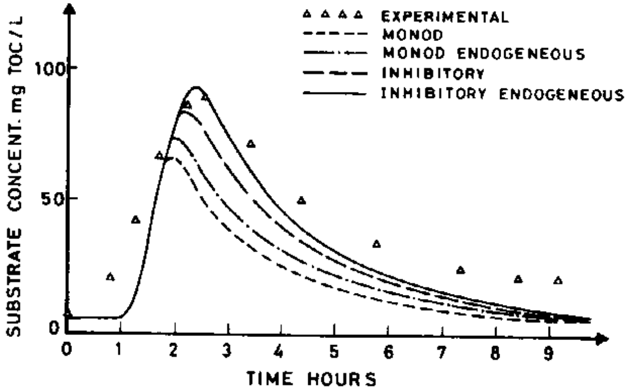


Fig. 17 . 10X Pulse transient stock (assuming a dead time of one hour and allowing for thickener dynamics)

c) The exact manner in which the pulse shock disturbance was imposed experimentally was not specified. This may explain the initial sharp increase in the simulated results since the model assumes that an ideal pulse disturbance was imposed.

d) The uncertainty regarding the actual aeration tank volume used (and hence τ) and recycle rate used. The technical report [5] from which the experimental data were taken indicates that the aeration tank volume can be varied from 3 to 8 litres by moving an internal partition. The effect of changing τ (as can be expected) has a significant effect on the value of the peak as well as the values of S achieved during the whole transient run. This is amply demonstrated in Fig. 18.

e) A microbial analysis of the biomass at different stages of the experimental results was reported in the technical report [5]. This indicated that certain cell population mutations occured and that filamentous fragments were observed in the photomicrographs. Since the four kinetic models we investigated assume that we are dealing with the same type of culture, such changes in microbial population and resulting kinetics were obviously not anticipated by the models.

f) The aeration chamber mixing character, location of feed and discharge and poor mixing can all have significant effects on the response.

7. Conclusions

It was shown that the Inhibitory and Endogenous metabolism model gave a better prediction of the transient response than that of the Monod model. Several

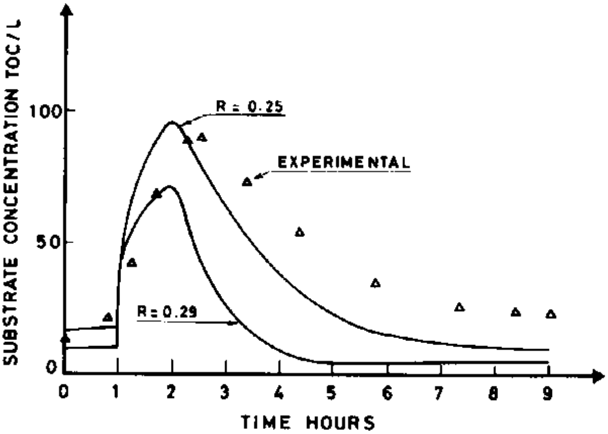


Fig. 18 . 10X Pulse transient (with thickener dynamics) τ = 4.4 hours

simplifications were made in the development of the models that resulted in limitations with regard to their applicatioṇ for the accurate prediction of the quantitative transient response of the activated sludge process. However, the simplicity of the model has the advantage of the relative ease of simulation to give relatively good qualitative results.

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تقويم بدائل نماذج حركية غير تركيبية لمحاكاة حركية عملية الحمأة المنشطة

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ملخص البحث: تم استخدام أربعة نماذج حركية غير تركيبية لمحاكاة الاستجابة الطارئة لوحدة الحمأة المنشطة التقليدية للتحميلات الصدمية. وقد افترض أن خزان التهوية هو عبارة عن مفاعل مقلّب مستمر، كما تمت محاكاة ديناميكية خزان الترسيب كتباطؤ بسيط من الدرجة الأولى. وقد تم اشتقاق معامل تركيز المواد العضوية والحيوية وذلك بالاستفادة من النتائج المختبرية المنشورة في هذا المجال. وقد كانت النتائج الخاصة بنموذج الإحياء الذاتي المانع جيدة النوعية بصورة نسبية. كما تمت مناقشة أسباب عدم ملاءمة هذه النماذج البسيطة في تقدير بعض الظواهر المهمة المتعلقة بالعمليات غير الثابتة.