OPTIMISATION AND UPRATING OF ACTIVATED SLUDGE PLANTS BY EFFICIENT PROCESS DESIGN

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ABSTRACT

Requirements to improve effluent quality and reduce operating costs at existing activated sludge plants in the UK have led to the development of an accurate mathematical model of the process which can be used for full-scale design. The WRc Activated Sludge Model has been developed over a number of years and is based on the concept of describing the kinetics of BOD removal by including a Monod term for growth and a Michaelis-Menten term for enzymatic activity. Since the first formulation of the model, further equations have been added which describe the use of oxygen and nitrate as electron acceptors for the conversion of BOD.

The model equations can be used to predict the performance of most configurations of the activated sludge process by using the 'tanks in series' concept to describe aeration tank mixing characteristics. Solutions to the equations are easily obtained by standard numerical integration techniques using a computer. The software has been written in such a way that the model can be used interactively by a plant designer.

Results obtained using the mathematical model have been used to redesign several full-scale activated sludge plants in the UK. In some cases, it has been possible to reduce the energy costs for aeration by 40% whilst maintaining effluent quality by accurately matching the supply of oxygen to the spatially-varying oxygen demand in plug-flow aeration tanks. Bulking sludge problems have also been solved by the conversion of completely-mixed aeration tanks to plug-flow aeration tanks with complementary redesign of the aeration system.

KEYWORDS

Aeration; Optimisation; Kinetics; Activated Sludge Model; Energy saving.

INTRODUCTION

It is well established that the activated sludge process can consistently produce effluents of high quality. However, the energy costs associated with the operation of activated sludge systems can be a high proportion of the total energy costs incurred for complete sewage treatment processes. Aeration is by far the largest energy consuming operation in the activated sludge process. Typically, the energy used in pumping recycled sludge is only approximately 5-10% of that used in aeration.

In the UK the effluent consent conditions applied to activated sludge plants usually mean in practice that the process must be designed to achieve carbonaceous oxidation only, or carbonaceous oxidation and nitrification of ammonia. The effluent quality criteria are approximately equivalent to mean values of BOD and suspended solids of 20 and 30 mg/l and ammonia concentrations of less than 5 mg/l respectively.
In recent years there has been considerable interest in reducing energy costs in the activated sludge process without affecting effluent quality. There is also a general trend towards improving effluent quality by uprating carbonaceous oxidation plants to provide completely nitrified effluents.

There are approximately 150 activated sludge plants in the UK which serve populations greater than 10,000. Of this number, approximately 40 installations employ fine-bubble aeration and approximately 100 utilise vertical-shaft mechanical aeration. In addition, there are a very small number of coarse-bubble aeration systems, seven 'Sheffield' paddle systems and a few large installations of the 'oxidation-ditch' configuration which employ horizontal-shaft aerators but which operate at conventional loadings. Exact analysis of site numbers is difficult since a few sewage works have more than one type of aeration system installed.

The size range of diffused-air activated sludge plants in the UK is quite different from the size distribution of mechanically-aerated systems. These distributions are shown in Fig. 1a and 1b.

There is only one diffused-air plant serving a population of less than 30,000. The four largest diffused-air plants treat over 55% of the sewage flow received by all diffused-air systems. The total equivalent populations served by diffused-air and surface aerated systems are almost identical at 11.2 and 11.5 millions respectively. However, over 50% of mechanically aerated systems serve populations of less than 40,000. There are only 10 sites serving populations in excess of 200,000.

The efficiency of the activated sludge process, in terms of energy used for aeration, can be determined by measuring the mass of oxygen transferred to the mixed liquor per unit of electrical energy expended. Thus the units of aeration efficiency could be kgO₂/kWh.

It is a very difficult procedure to measure aeration efficiency in an existing activated sludge system without disturbing normal effluent quality. It is convenient to use a simple, but fairly inaccurate, method which calculates aeration efficiency indirectly by estimating the mass of oxygen required to achieve the degree of treatment obtained in a given system. In general, oxygen is required to satisfy BOD, to maintain the microbial population and, if necessary, to achieve nitrification. It should, therefore, be possible to relate the rate of oxygen transfer to the mixed liquor to the flowrate and composition of the sewage and to the effluent quality obtained.

The following equation can be used to assess the oxygen demand of a given activated sludge process with reasonable accuracy:

\[
\text{Oxygen demand} = 0.0864q_s \left[0.75(BOD_i - BOD_e) + \frac{5.25 \times 10^{-4} \times \text{MLSS} \times V}{q_s} + 4.3(N_i - N_e)\right] \quad (\text{kg/d}) \tag{1}
\]

Where:

- \(q_s\) = Average settled sewage flowrate (l/s)
- \(BOD_i, BOD_e\) = Average BOD concentrations in settled sewage and effluent, respectively (mg/l)
- \(\text{MLSS}\) = Mean MLSS concentration (mg/l)
- \(V\) = Aeration tank volume (m³)
- \(N_i, N_e\) = Average ammonia-nitrogen concentrations in settled sewage and effluent respectively (mg/l)

Equation (1) is not sufficiently accurate for use as a standard performance test of aeration equipment. However, it is useful for comparative assessment of the aeration efficiencies of operational activated sludge plants. The oxygen demand value obtained from Eq. (1) can be converted into an aeration efficiency by dividing by the daily consumption of electrical energy used in aeration. (kWh/d).

Houck and Boon (1981) used Eq (1) to calculate the aeration efficiencies of severaldiffused-air activated sludge plants. The results indicated that there was a wide variation in efficiencies between plants producing effluents of comparable quality. The aeration efficiencies of 13 plants investigated were in the range 1.1-2.2 kgO₂/kWh.

In the UK, variation in aeration efficiency from site to site is largely a function of differences in process design and is not particularly dependent on the type of
Figure 1a Size Distribution of Diffused Air Plants

Figure 1b Size Distribution of Surface Aeration Plants
aeration equipment installed. However, there is evidence that fine-bubble, diffused-air systems can only realise their maximum aeration efficiency in fully-nitrifying processes. In highly-loaded, carbonaceous treatment processes diffuser blockages due to internal slime growth can be a problem. In the UK, nearly all the diffused-air installations produce nitrified effluents and surface aeration systems are usually installed in highly-loaded processes.

Improvements in aeration efficiency and hence potential savings in energy, can be achieved by giving attention to the following aspects of process design:—

Matching Oxygen Supply to Oxygen Demand

In many operational activated sludge systems it is often observed that aeration tank dissolved oxygen concentrations tend to increase towards saturation values as the tank outlet is approached. There are two main reasons for this effect,

(a) the aeration tank may be severely underloaded, either by deliberate conservative design or by operational policy, and

(b) the aeration equipment may not have been installed to meet the spatially-varying oxygen demands which occur in most aeration tanks.

Deliberate overdesign is reasonable in order to guarantee effluent quality if no precise design procedure is available. In the past, the cost penalty of inefficient energy usage has been acceptable.

Fine-bubble diffused-air systems have traditionally been installed in aeration tanks which tended to exhibit plug-flow mixing characteristics. The advantage of tapered-aeration systems, where the number of diffusers is gradually reduced from inlet to outlet, has long been recognised. The variation of oxygen demand in an aeration tank is a complex function of tank geometry and treatment capacity. Previously available design procedures have resulted in ‘standard’ tapered-aeration diffuser layouts being installed in widely different situations. The result has been a wide range of operational aeration efficiencies.

In mechanical surface-aeration systems attempts at tapered aeration have been made by installing aerators of different power rating in multi-compartment systems. There can be constraints on this procedure since aerator sizes must be within certain well defined ranges for a given size of aeration tank.

Installation of a DO Control System

The advantages of DO control in activated sludge aeration tanks have been recognised for many years, both from the energy saving and effluent quality standpoints. The availability of robust, easily maintained DO probes has removed previous difficulties encountered with earlier, less reliable instruments. For maximum effect, a DO control system must be installed in conjunction with a properly designed aeration system. In some diffused-air processes the number of diffusers is so excessive for the load applied that the air flowrate is already at the minimum permissible for each diffuser. High aeration efficiencies and consistent effluent quality can be achieved by optimising the layout of aeration equipment to meet the average load conditions applied and then installing a DO control system to vary the aeration intensity as required during periods of diurnal variation.

The various types of aeration equipment available have different turndown capabilities and there are limits to the maximum range of oxygenation capacity which it is possible to achieve in practice.

Experience has suggested that maximum efficiency may not result from the use of one DO probe to control the output of the entire aeration equipment in an aeration tank. It may be necessary to divide the aeration system into two or more independently controllable zones, each with an associated DO probe and DO set-point.

It might be considered that the difficulties previously mentioned could be overcome easily by designing all activated sludge aeration tanks on a completely-mixed basis. Under such conditions, there would be no spatially-varying oxygen demand and the installation of optimised aeration and DO control systems would be greatly facilitated. Unfortunately, it has been demonstrated by a number of workers that
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completely-mixed systems are much more likely to suffer from bulking sludge than activated sludge processes designed on a plug-flow basis. Chudoba et al. (1973), Rensink et al. (1982) and Chambers (1982).

The basis of most process engineering design procedures is a mathematical model which accurately describes the behaviour of the systems in terms of the fundamental variables of interest. The design of efficient activated sludge systems requires a model which is accurate enough to permit traditional safety margins to be considerably reduced with confidence. Firstly, the model must describe the complex microbial reactions which occur in activated sludge in terms of variables which can be easily measured as part of routine analysis. Secondly, the model must be applicable to a wide range of operating conditions and aeration tank mixing regimes and, thirdly, the complexity of the model should be balanced against its ease of use.

The next sections of this paper describe the development and formulation of a mathematical model of the activated sludge process and give examples of its use in the optimisation and uprating of full-scale plants.

WRc ACTIVATED SLUDGE MODEL

Introduction

The model was first published by Jones (1973). Work carried out at the Water Pollution Research Laboratory had demonstrated the difficulty of reconciling the biochemical activity of activated sludge with the numbers of viable bacteria present. In some cases, the observable potential biochemical activity was an order of magnitude greater than could be accounted for by the concentration of viable cells in the sludge. Further, in considering the energetics of the process, calculation showed that for a typical settled sewage the quantity of assimilable carbon present could maintain no more than 20% of the sludge as viable bacteria.

Kinetics of waste-water treatment

Downing et al. (1964) had developed a model, based on Monod kinetics, (Monod 1942, 1950), which accurately described the performance of the activated sludge process with respect to nitrification. In the years that followed, several other models emerged, (Downing and Knowles 1967; Curds 1973) applying the same, or similar, equations to protozoa and to heterotrophic bacteria. While these models were often adequate when applied to completely-mixed aeration systems, problems arose with the degree of treatment predicted, particularly as longitudinal mixing was decreased, and also with the predicted concentration of viable bacteria within the system. Values for $K_s$ ($K_s =$ the concentration of the substrate at which the specific growth is half the maximum specific growth rate) in the range of 100-200 mg BOD/l had been reported by Tench and Morton (1962); many models used similar values for $K_s$ in order to overcome the prediction of a vanishingly small concentration of substrate in the final effluent.

Viability

Monod developed his kinetic equations on the basis of bacterial growth in laboratory fermenters in which viability of the population was almost 100%. The conditions in an activated sludge plant are very different. Firstly, the concentration of the feedstock entering the plant is very low (Painter et al., 1961), and secondly the object of the treatment process is to reduce, as far as possible, the concentration of metabolisable substrate. This makes the conditions in a treatment plant more akin to those in a spent batch culture, where the viability of the culture is beginning to decline. Although a bacterium may be 'non-viable', this does not necessarily mean that it is no longer capable of biochemical activity; the activity may even be enhanced as was shown by Stephenson (1928) for lactic dehydrogenase.

Wooldridge and Standfast (1933, 1936) recognised that the majority of the bacterial population in activated sludge was non-viable. Their experiments showed conclusively that non-viable bacteria were quite capable of removing Biochemical Oxygen Demand (BOD) from a reaction mixture in the absence of any viable cells. Indeed, they concluded that since there was such a large proportion of non-viable cells in activated sludge, these cells would need to retain only a part of their total activity for them to be responsible for the majority of the biochemical activity observed in the sludge. Results from more recent chemostat studies are shown in Table 1 which
demonstrate that viability of the population does not need to be 100% in order for full biochemical activity to be exerted when the growth substrate is added to a washed suspension of the harvested cells.

**TABLE 1**

<table>
<thead>
<tr>
<th>CULTURE VIABILITY</th>
<th>RESPIRATION RATE</th>
<th>INCREASE IN RESPIRATION DUE TO GLUCOSE</th>
</tr>
</thead>
<tbody>
<tr>
<td>%</td>
<td>mgO₂/gcells.hr</td>
<td></td>
</tr>
<tr>
<td>40</td>
<td>56</td>
<td>194</td>
</tr>
<tr>
<td>100</td>
<td>24</td>
<td>153</td>
</tr>
</tbody>
</table>

Description of model

The model consists of a number of first-order differential equations derived from a mass balance around a completely-mixed reactor. Provision is made to simulate from 1 to 12 completely mixed tanks in series, providing hydraulic regimes from completely mixed to pseudo plug-flow. Also incorporated is a rudimentary model of the final sedimentation tank.

Each compartment contains an identical number of constituents with a differential equation for each. These constituents are BOD, viable heterotrophs, non-viable heterotrophs, ammonia, viable nitrifiers, non-viable nitrifiers, total solids, nitrate and dissolved oxygen. The equations are solved simultaneously with a variable step fourth-order Runge Kutta numerical integration routine.

**BOD and Ammonia**

The model recognises that utilisation of substrate can occur within the process without this consumption being coupled to growth. Thus a more accurate description of the removal of substrate from a waste-water treatment system would include in the mass balance equation both a Monod term for the conversion of substrate to new biomass, and a Michaelis-Menten term for consumption of substrate without the formation of biomass. Equation (2) shows the mass balance for a single completely-mixed reactor with recycle.

\[
\begin{align*}
\frac{ds}{dt} &= q \left( S_0 + r S_r \right) - \frac{\mu m_y}{m_y} S - \frac{r C_n}{C_n} \frac{S}{K_m + S} - q(1+r)S \\
&= q \left( S + rS_r \right) - \frac{\mu m_y}{m_y} S - \frac{r C_n}{C_n} \frac{S}{K_m + S} - q(1+r)S \\
&= \frac{Y(K + S)}{K + S} \frac{S}{K_m + S} \frac{V}{m_y} \text{ MONOD MICHAELIS-MENTEN} \\
\end{align*}
\]

Where:-
- \( q \) = Influent flowrate
- \( V \) = Reactor volume
- \( S \) = Influent concentration of growth limiting substrate
- \( S_0 \) = Concentration of limiting substrate in recycle
- \( r \) = Recycle ratio
- \( \mu m \) = Maximum specific growth rate
- \( C_v \) = Concentration of viable cells
- \( Y \) = Yield co-efficient (g cells produced/g substrate produced)
- \( K_m \) = Concentration of growth limiting substrate, \( S \), at which the specific growth rate of the viable cells is half the maximum activity
- \( C_n \) = Concentration of non-viable cells

The same equation, with the appropriate kinetic constants, is used for both BOD and ammonia. In the case of oxidation of ammonia the model assumes a stoichiometric conversion to nitrate.
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Biomass

An essential part of the model is the transition from 'viable' to 'non-viable' cells. Heterotrophic micro-organisms isolated from sewage treatment plants have been found to have maximum specific growth rates of about 0.3h⁻¹ at 15 to 20°C. This compares with specific growth rates of 0.008h⁻¹ and less for typical activated sludge plants. At such low proportions of their maximum specific growth rate bacterial cultures are known to lose viability, and the probability of cell division resulting in the production of a non-viable cell increases, (Tempest et al., 1987). Experiments at WPRL showed that at 10% of the maximum specific growth rate of the culture there was a marked decline in the viability of that culture (WPRL, 1973). For the purposes of the model it is assumed that the organisms remain fully viable while the concentration of the growth limiting substrate is able to support at least 10% of the maximum specific growth rate. The function used to calculate the viability (maximum value = 1.0) is show in Eq.3.

\[
\text{Viability} = \frac{S}{f(K_s+S)} \tag{3}
\]

Where

- \( f \) = the proportion of the maximum specific growth rate below which the cells lose viability (assumed usually to be 0.1)
- \( K_s \) = the concentration of the growth limiting substrate, \( S \), at which the specific growth rate is half the maximum.

The viability of the population returned to the aeration tank is calculated using the concentration of the substrate in the underflow from the final sedimentation tank.

In addition, there is assumed to be an endogenous decay of solids, and the mass balance for solids around a completely mixed tank is given in Eq.4.

\[
\frac{dX}{dt} = \frac{q}{V} (X_o + rX_r) + \mu c_v - kX - \frac{qX(1+r)}{V} \tag{4}
\]

Where

- \( q \) = Influent flowrate
- \( V \) = aeration tank volume
- \( X \) = the concentration of solids in the influent
- \( X_o \) = concentration of solids in the recycle
- \( r \) = recycle ratio
- \( \mu \) = specific growth rate
- \( C_v \) = concentration of viable cells
- \( k \) = specific rate of decay of the solids
- \( X \) = concentration of solids in the aeration tank.

Utilisation of Oxygen

As an aerobic process, the treatment of a waste-water by activated sludge is influenced by the partial pressure of oxygen within the bulk of the liquid. For the purposes of the model, oxygen is considered as another substrate, consumed concurrently with either carbonaceous BOD or ammonia, which moderates both the Monod and Michaelis-Menten terms in Eq.2 according to the relationship,

\[
\text{Rate} = \frac{pO_2}{K_o + pO_2} \tag{5}
\]

Where \( pO_2 \) = partial pressure of oxygen within the bulk of the liquid
- \( K_o \) = saturation constant for oxygen at which the reaction rate is half the maximum.

For simplicity in the calculations, and presentation of the results of the model, concentration of oxygen is expressed in terms of mg dissolved oxygen/litre and an atmospheric pressure of 760 mm of mercury is assumed. The values of \( K \) for carbonaceous BOD and ammonia are taken to be 0.1 and 1.0 mg dissolved oxygen/l respectively. Additionally, the model allows for a further oxygen demand, representing 'endogenous respiration', which is also moderated according to the
relationship shown in Eq. 4, using a value for $K_o$ of 0.1 mg/l.

The rate of solution of oxygen in the model is calculated using the expression

$$\text{Rate of solution} = k_L a (C_s - C_t)$$

Where $C_s$ = saturation value for dissolved oxygen
$C_t$ = concentration of dissolved oxygen
$k_L a$ = oxygen transfer rate.

At initialisation the saturation value for dissolved oxygen is calculated from the specified temperature. The maximum rate for $k_L a$ is one of the parameters provided by the user. When dissolved oxygen control is requested, intermediate values for $k_L a$ are calculated to maintain the concentration of dissolved oxygen at the required level.

**Denitrification**

Activated sludge contains bacteria able to use nitrate ion as an alternative electron acceptor to oxygen thereby reducing nitrate, via nitrite, to gaseous nitrogen. This switch to an alternative electron acceptor occurs when the rate of reaction supportable by the nitrate ion is greater than that which can be supported by the dissolved oxygen present (Paskins et al., 1978). Anoxic zones are exploited in the activated sludge process to effect an overall reduction in the inorganic nitrogen discharged in the final effluent. The equations included in the model, which allow the specification of anoxic zones, are identical to those used for normal aerobic metabolism, using different kinetic constants. When an anoxic zone is specified, the specific growth rate that can be achieved with oxygen is compared with that attainable using nitrate, and the greater of the two is then used in the integration.

**Sedimentation**

Separation and recycle of the solids is the essential feature of the activated sludge process. Recycle of the biomass depresses the specific growth rate below the dilution rate of the aeration basin, and enforces the characteristic ‘old batch culture’ conditions necessary for the attainment of a highly treated effluent. If a model is to be used to make realistic simulations of the dynamic behaviour of the system then it is also essential that an adequate delay in the return of the solids from the outlet to the inlet of the aeration tank be included.

The model of the final sedimentation tank is based on the work of White (1975). Four compartments, of equal volume, in series, are used to simulate the body of the tank. The first compartment represents the upper part of the tank containing the settled effluent, the second compartment receives the mixed liquor from the aeration stage, the third and fourth compartments represent the body of the tank, the recycled sludge being drawn from the fourth compartment. Total volume of the sedimentation tank is calculated from the user specified surface area of the tank, with minimum and maximum depths of 2.0 and 3.0 metres respectively. Solids are transported through the tank both hydraulically and by gravity. The settling velocity of the sludge across the boundary between two compartments due to gravity is determined by an equation of the form,

$$v = v_o e^{-kc}$$

Where $v$ = maximum rate of settlement
$k$ = related to the settleability of the sludge
$c$ = concentration of solids in the lower compartment.

**Description of aeration tank longitudinal mixing**

The degree of longitudinal mixing in an aeration tank can be determined by a tracer test. Parameters used to describe longitudinal mixing in activated sludge plants can be related to sludge settleability as shown in Figure 2. This curve refers to a survey of plants in the UK and is accurate enough to be used for design purposes. An obvious difficulty with tracer tests is that the process must exist before the test can be performed. The parameter used in Fig. 2 to quantify longitudinal mixing is the dimensionless Dispersion Number $D$ (Levenspiel, 1962). Examination of Fig. 2 indicates that it would be advantageous to design aeration tanks with values of the Dispersion Number of less than approximately 0.06. The problem then arises of how to relate Dispersion Numbers to the geometry and operating conditions of the aeration tank.
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The variables which are combined in the Dispersion Number are: D, a dispersion co-efficient having the dimensions of m$^2$/s, $u$, the average fluid velocity in the aeration tank; and $L$, the total aeration tank length. The Dispersion Number may be rewritten as:

$$\text{Dispersion Number} = \frac{DWH}{Lq_s(1+r)}$$  \hspace{1cm} (8)

Where $D$ = Dispersion Co-efficient (m$^2$/s)
$W$ = Aeration tank width (m)
$H$ = Aeration tank depth (m)
$L$ = Aeration tank length (m)
$q_s$ = Average sewage flowrate (m$^3$/s)
$r$ = Recycle ratio (-)

Eq. 8 relates the aeration tank Dispersion Number to the tank geometry and operating conditions. The Dispersion co-efficient, D, can be considered to be a measure of the magnitude of turbulent eddies in the aeration tank. In diffused-air systems $D$ will be largely influenced by the rate of energy dissipation per unit volume of aeration tank. Measurements of Dispersion Numbers performed by tracer tests in full-scale plant have revealed that the value of $D$ can be considered virtually constant in diffused-air systems and has a value approximately equal to 0.068 m$^2$/s. Chambers and Thomas (1985). Eq. 8 then becomes:

$$\text{Dispersion Number} = \frac{0.068 WH}{Lq_s(1+r)}$$  \hspace{1cm} (9)

Eq. 9 can be used to estimate the degree of longitudinal mixing in aeration tanks with an accuracy of ±15%, for diffused-air systems only, providing the variables are within the following ranges:

$$2 < W < 20 \text{ (m)}$$
$$2.4 < H < 6.0 \text{ (m)}$$
$$28 < L < 500 \text{ (m)}$$
$$0.7 < r < 1.5 \text{ (-)}$$
$$1.3 < t < 8.0 \text{ (h)}$$

Where $t = \frac{V}{3600q_s(1+r)}$ and $V$ = aeration tank volume (m$^3$)

The Dispersion Number approach can be used to generate a realistic set of equations which describe the behaviour of activated sludge processes. Olsson and Andrews (1978) have shown how such a model can be used to predict the dissolved oxygen profile in an...
aeration tank. However, the description of non-ideal flow behaviour using the Dispersion Number concept leads to quite complex numerical techniques being required for solution of the model equations.

An alternative approach to the description of longitudinal mixing is the so-called 'tanks in series' model. The main advantage of this model is that quite complex models of the activated sludge process can be developed without generating differential equations higher than first order. The basis of the tanks in series approach is the assumption that real flow systems are equivalent to an appropriate number of completely mixed tanks connected in series. Solution of model equations is facilitated since discontinuities in parameter values occur only at the boundaries between theoretical tanks.

The 'tanks in series' model is obviously appropriate for aeration tanks which are physically divided into compartments, such as surface-aeration systems. The appropriate value of 'N', the number of tanks in series, should be obvious from visual inspection of the system and this value can be used in simulation of the process.

The tanks in series model can also be used to describe longitudinal mixing in flow systems which are not physically divided into real compartments. Although no absolute comparison between the tanks in series and Dispersion Number models exists, it is sufficiently accurate to assume the following simple relationship:

\[ \text{Dispersion Number} = \frac{1}{2N} \]  

Combining Eq. 9 and 10 gives the following equation:

\[ N = \frac{7.4 \times Lq \times (1+r)}{WH} \]  

The WRc model of the activated sludge process assumes that longitudinal mixing can be described by an appropriate value of 'N', the number of tanks in series to which the aeration tank is equivalent. A representation of the activated sludge process as used in the model is given in Fig. 3. The model equations consist of a series of mass balances on various components in the system. A mass balance on dissolved oxygen for an arbitrary jth tank in the system can be written:

Accumulation of oxygen in tank = Bulk flow of oxygen into tank - Bulk flow of oxygen leaving tank + oxygen transferred to liquid from air - oxygen consumed by micro-organisms

In symbols the equation becomes:

\[ \frac{V}{N} \frac{dC_j}{dt} = q_s (1+r) C_{j-1} - q_s (1+r) C_j + (k_{L,a})_j \frac{V}{N} (C_s - C_j) - R_j \]  

Where \( C_j \) = Concentration of DO in jth tank  
\( C_{j-1} \) = Concentration of DO in (j-1)th tank  
\( (k_{L,a})_j \) = Overall oxygen mass transfer co-efficient for oxygen in jth tank  
\( R_j \) = Rate of consumption of DO by micro-organisms.

The rate of consumption of oxygen is a function of the variables previously described.

\[ \frac{V}{N} \frac{dC_j}{dt} = q_s (1+r) C_{j-1} - q_s (1+r) C_j + (k_{L,a})_j \frac{V}{N} (C_s - C_j) - R_j \]  

Figure 3. Representation of Activated Sludge Model
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Solution of the model equations is a relatively simple procedure using numerical methods on a computer. The program has been written in such a way that the process is simulated on the screen of a computer terminal. Eq. 12 can be solved for any integer value of N between 1 and 12. A time variation in sewage flowrate and organic concentration can be used to simulate diurnal inputs to the system.

USE OF THE ACTIVATED SLUDGE MODEL FOR OPTIMISATION AND UPRATING OF FULL-SCALE PLANT.

The activated sludge model has been used to optimise the design and performance of new and existing treatment plants at several sites.

At the Rye Meads STW of Thames Water, the aeration efficiency of a nitrifying activated sludge plant was increased from 1.2 kg O₂/kWh to 1.8 kg O₂/kWh by the installation of an optimised fine-bubble diffuser system and by efficient DO control. Full details of this project have been reported in the literature (Chambers et al., 1985; Robertson et al., 1984).

Another full-scale optimisation project is nearing completion at the Blackburn Meadows STW of Yorkshire Water. A large secondary treatment system employing vertical-shaft mechanical aerators to produce a non-nitrified effluent has been re-designed. Results of simulation exercises were used to predict aeration power requirements and an optimised aeration system was installed. The results of several months of operation have been reported (Thomas and Chambers, 1982).

REFERENCES


