A UNIFIED APPROACH TO THE DESIGN AND OPERATION OF THE ACTIVATED SLUDGE SYSTEM

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INTRODUCTION

Advancements made by Dick [1-3] and Vesilind [4] with respect to the solids thickening function of final clarifiers in the activated sludge treatment system served to transform existing system design methodologies to a rational basis. Their developments, based on the theoretical contributions of Coe and Clevenger [5], Kynch [6] and Yoshioka, *et al.* [7], considered system interactions between the aeration basin and final clarifier. The first objective of this chapter is to demonstrate how these developments, herein termed the settling flux approach, can be adapted for evaluating economic trade-offs between alternative system designs and for establishing least-cost designs.

Although the settling flux approach is now occasionally employed for system design, the technique has not yet been applied to system operation. Accordingly, the second objective is to demonstrate that the settling flux approach also can be employed by treatment plant operators to monitor the operational state of an activated sludge system. Biosolids inventory control decisions can thereby be made to (a) prevent system failure, (b) to maximize overall treatment efficiency and (c) to minimize energy consumption and the resultant costs of operation. It will be shown that the settling flux approach is effective for making control decisions in response to both short-term (diurnal) and long-term (seasonal or yearly) changes in the plant influent and system operational conditions.

While the unified systems approach, developed for designing and operating an activated sludge treatment train, has been adapted for computer solution, only the methodology is presented herein. This permits the reader the opportunity to review the fundamental considerations and to appreciate the simplicity and versatility of this approach.

BACKGROUND

Only a brief introduction to the theory of thickening is included herein. For a comprehensive treatment, the reader is referred to Dick [8] or Vesilind [9]. The basic activated sludge system considered in this development is shown diagrammatically in Figure 1.

Biosolids Sedimentation

Biosolids that are introduced into a clarifier are transported to the bottom of two velocity components (see Figure 2): v, the gravitational sedimentation component and

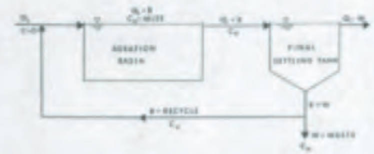

Figure 1. Schematic of activated sludge system.

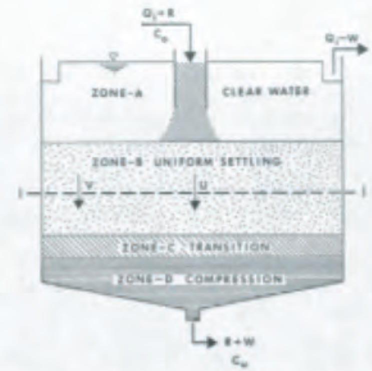

Figure 2. Cross section of secondary clarifier.

u, the component due to withdrawal of solids in the underflow (u = R/A). The solids transport flux due to the former, termed settling flux, is defined as the product of the solids concentration and settling velocity.

$$G_s = cv \tag{1}$$

The solids flux due to the latter, termed the bulk flux, is equivalent to the product of the solids concentration and the bulk underflow withdrawal velocity

$$G_b = cu \tag{2}$$

The total transport flux, therefore, is given by the sum of the two components

$$G = G_s + G_b \tag{3}$$

All parameters are defined in the section entitled Notation.

Schematically, the two flux components and the total flux can be represented as shown in Figures 3 and 4, respectively. A minimum in the total flux curve can be observed in Figue 4. This, termed the limiting flux, G_L, is the maximum solids flux that can be transmitted to the bottom of the clarifier. The limiting flux establishes the solids-handling capacity of the clarifier.

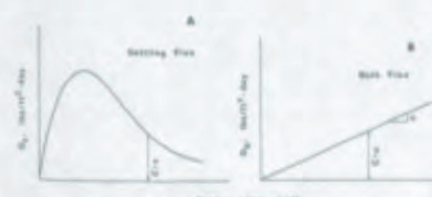

Figure 3. Settling and bulk flux solids transport components.

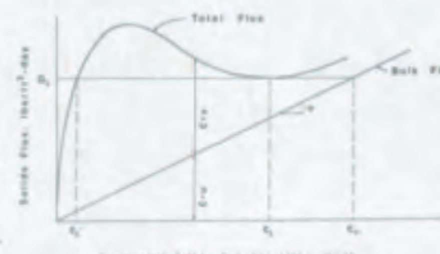

Figure 4. Total flux curve for sedimentation.

Batch Settling Flux Approach

A geometric technique developed by Yoshioka *et al.* [7] allows one to determine the limiting flux simply by drawing a line tangent to the settling flux curve which intersects the s-axis at the desired value of underflow solids concentration, C_u (Figure 5). This line has a slope of minus u. Similarly, a line drawn from the origin to any point on the settling flux curve has slope of v. This is the settling velocity of the solids at the concentration that corresponds to the x-coordinate of the point of intersection.

Figure 6 illustrates the analogy between the total flux curve (Figure 4) and the geometric technique developed by Yoshioka *et al.* [7]. The latter approach has been employed in this development because it offers the advantages of simplicity and versatility of application.

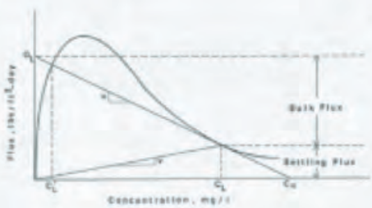

Figure 5. Batch settling flux analytical method.

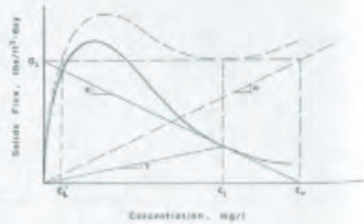

Figure 6. Batch and total settling flux curves.

Solids Concentration Profiles

Solids concentration profiles that exist in operating prototype secondary clarifiers can be approximated using the batch settling flux curve. This requires the construction of an operating line which has a Y-intercept equal to the applied flux, $C_o(Q_i + R)/A$, and a slope of u (R/A). As noted above, the X-intercept should be equal to the suspended solids concentration in the underflow of the clarifier, C_u.

Three illustrative operating lines have been constructed on the batch settling flux plot shown in Figure 7. The operating line for Case A is shown tangent to the batch flux curve. For such a condition the clarifier is critically loaded; *i.e.*, the mass flow of solids entering the clarifier is equal to the maximum rate at which solids can be transmitted to the bottom. For this case one would expect to observe three stable suspended solids "blankets" of different concentrations. Figure 8 shows the expected concentration-versus-depth profile. The three stable concentrations shown, C'_L, C_L and C_u correspond to the

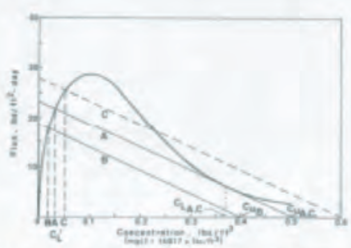

Figure 7. Clarifier analysis for three loading cases.

Figure 8. Solids concentration profiles.

concentrations defined by the intersection of the operating line with the rising limb of the batch flux curve, the point of tangency of the operating line and the batch flux curve and the intersection of the operating line and the x-axis, respectively.

The upper dilute blanket, C'_L, would be expected to extend downward from the clarifier feed point to the thick blanket, C_L. The upper dilute blanket, however, may not form if the solids are flocculant. In such cases the solids would fall from the feed point as an "underwater waterfall" to the point at which they are incorporated into the thick blanket as illustrated in Figure 2. The zone of concentration C_u would exist only near the bottom of the clarifier.

In an underloaded clarifier, Case B, only two stable blankets would be expected to form: the upper dilute blanket of concentration C'_L and the underflow zone of concentration C_u. Both C'_L and C_u are less than the corresponding concentrations for Case A (see Figures 7 and 8).

An overloaded clarifier, Case C, is described by the dashed operating line in Figure 7 which passes above the batch settling flux curve. Because the clarifier is able to transmit only the limiting flux to the bottom, all solids applied to the clarifier in excess of the limiting flux must accumulate within the clarifier. This occurs by the upward propagation of the blanket of concentration C_L. Ultimately the blanket propagates upward to the point where the excess solids are scoured from the blanket and are discharged over the weirs of the clarifier. In this case the upper dilute blanket of concentration C'_L occurs only during the period that the thick blanket is below the feed point. It is to be noted that the concentrations C_L and C_u are the same as those that would be observed in the critically loaded case.

Although the concentration profiles shown in Figure 8 have been illustrated as being discontinuous ("squared") in shape, the actual profiles generally are considerably rounded; that is, the concentration gradients are not as pronounced as illustrated. This occurs because of hydraulic solids dispersion in the clarifier.

State Point Concept

For design and operations analysis it is desirable to introduce the concept of a state point, first defined by McHarg [10]. The state point is defined by the intersection of two operating lines on the batch flux plot. One of the lines has a slope of minus u and describes the recycle flow from the clarifier to the aerator (Figure 9). The other begins

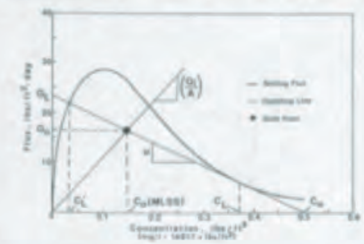

Figure 9. Batch settling flux curve with operating lines and state point.

at the origin and has a slope equivalent to the clarifier overflow rate, Q_i/A, expressed in feet per day. (Actually the clarifier overflow rate is equal to $(Q_i - W)/A$. Since W is usually less than 3% of Q_i, the overflow rate can be approximated by Q_i/A without appreciable error). Accordingly, the state point has a y-coordinate of G_o, which is the solids flux, C_oQ_i/A, imposed on the clarifier due to influent flow and an x-coordinate of C_o, the mixed liquor suspended solids concentration. It must be emphasized that C_o is not intrinsically a concentration that one would expect to observe in the clarifier. Nonetheless, either the dilute or thick blankets could have a concentration equal to C_o under certain operating conditions.

To illsutrate the utility of the state point concept two examples are instructive. Figure 10 shows that the state point serves as a pivot for the situation in which the recycle flow rate is changed. By decreasing the recycle flow rate the slope of the recycle rate operating line also decreases proportionately; *i.e.*, $u_4 < u_3 < u_2 < u_1$. It is to be noted,

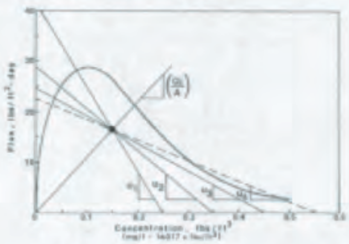

Figure 10. Batch settling flux curve with four recycle rate operating lines.

furthermore, that the dashed recycle rate operating line with slope u_4 passes above the batch settling flux curve. This indicates that recycle flow rate has been decreased to the extent that the clarifier is overloaded with respect to its thickening capacity. This would result in the transfer of biosolids from the aerator to the clarifier. These solids would be sorted in the clarifier in the form of a solids "blanket" of approximate concentration C_L. The blanket would propagate upward in the clarifier to accommodate the solids transferred. If the overloaded condition prevailed for an extended period of time, the clarifier would ultimately fail by scouring solids from the blanket and discharging them over the weirs. This overload condition could be relieved by increasing the recycle rate such that the recycle operating line does not pass above the settling flux curve.

Diurnal changes in the influent flow rate to the activated sludge system change the location of the state point as may be observed in Figure 11. Shown are influent overflow rate operating lines for three cases of diurnal flow rate: $(Q_i/A)_{max}$, $(Q_i/A)_{avg}$ and $(Q_i/A)_{min}$. All three recycle rate operating lines are parallel to one another because the

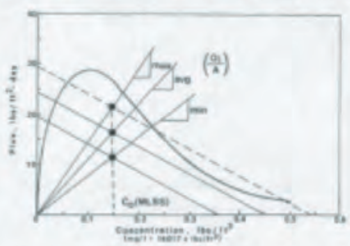

Figure 11. Batch settling flux curve with three overflow rate operating lines.

recycle flow rate was not changed. The state point is seen to shift from one (Q_i/A) operating line to another at constant C_o (MLSS). An increase in the influent flow rate to the system does not *a priori* result in a redistribution of biosolids between the aerator and clarifier.

As in the previous case, the dashed recycle rate operating line passes above the batch flux settling curve, indicating an overloaded condition. This could be resolved, as before, by increasing the recycle flow rate such that the recycle operating line is rotated to the point of tangency. Otherwise some biosolids would be transferred from the aerator to the clarifier causing the thick "blanket" to rise in the clarifier while the solids concentration in the aerator becomes more dilute.

Experimental Approach and Apparatus

Use of the batch flux approach requires experimental development of settling flux curves. This is accomplished by determining the initial interface settling velocities (ISV's) of a representative sludge sample at different initial concentrations. Interface settling curves that were obtained for a sludge obtained from an extended-aeration activated sludge treatment plant have been plotted in Figure 12 for initial suspended solids concentrations ranging from 3122 to 13,497 mg/l. Initial settling velocities are determined from the slope of the linear segment of the traces. In doing so it is important to exclude the curved portion of the traces where the settling velocity of the sludge increased due to reflocculation.

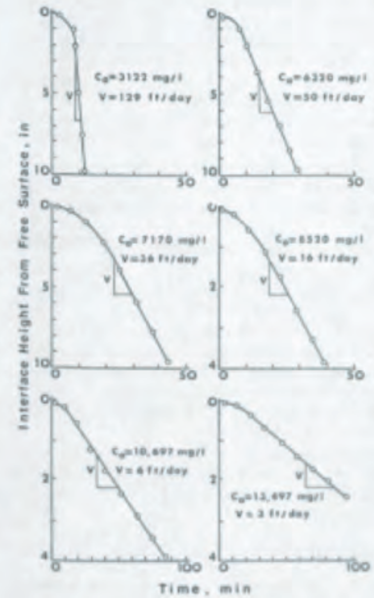

Figure 12. Solids/liquid interface height versus time traces.

Experimental interface settling velocities are subsequently plotted as a function of sludge solids concentration (Figure 13A). Since the settling flux is defined as the product of the solids concentration and the corresponding initial interface settling velocity, the settling flux curve (Figure 13B) is developed by plotting the product of the ordinate and abscissa

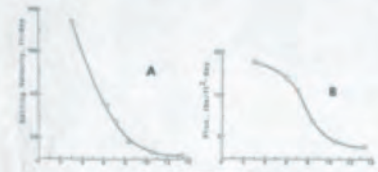

Figure 13. ISV and settling flux versus concentration.

values of the settling velocity versus solids concentration curve as a function of solids concentration. The resultant batch settling flux curve serves as the basic working diagram for all design and operational control procedures.

It must be emphasized that selection of a representative sludge sample is of paramount importance in developing the batch settling flux curve. For design purposes, a representative sludge may be obtained either from an existing facility that is operated under conditions that are similar to those specified for the plant which is being designed or from a pilot plant when no representative sludge is otherwise available. Because sludge settling characteristics change with solids residence time, seasonal temperature and biomass population shifts, among other factors, all such factors must be considered when selecting a representative sludge.

When analyzing the operational state and making operations control decisions for an existing activated sludge system, a representative sludge is available at all times. The question that remains, however, is "how frequently should the gravitational settling velocity tests be performed to ensure that the settling flux curve employed for analysis is, indeed, current and applicable?" No specific guidance can be given in this regard at this time, since changes in sludge settling characteristics are system-specific and because no quantitative information has yet been elucidated. Current research efforts are being directed toward this end–that of defining the dynamics of sludge settling characteristics.

To obtain laboratory interface settling velocities that are representative of settling velocities that occur in prototype secondary clarifiers, the following conditions outlined by Vesilind [9] must be observed when constructing the required laboratory equipment and when conducting the tests themselves:

1. The settling-column diameter should be as large as possible–8 in. is a practical compromise.
2. The initial liquid depth should be the same as the depth of the prototype thickener. When this is impracticable, 3 ft should be considered as the minimum.
3. The sludge slurry should be slowly introduced into the column through the bottom to minimize turbulence and the attendant shearing of floc.
4. The sample should be slowly stirred throughout the test using a "rake" type of mechanism. Reasonable rotational speeds range from 0.5 to 1.0 rpm.
5. The temperature of the sample must be maintained constant throughout the test, preferably within a 3-degree (C) range.

An analog/digital solids-liquid interface settling monitor developed by George [11] and shown in Figure 14 meets the above criteria and provides for the automatic off-line measurement of the batch settling properties of biosolids. Alternatively, the solids-liquid interface could be monitored manually.

SYSTEM DESIGN

Interactions among the various components of wastewater treatment systems must be recognized if cost-effective designs are to be achieved. In most previous designs of the aeration basin/clarifier system, the two units have been treated as distinct entities that have no interactions. Dick [2] has illustrated the interactions between the two processes and has developed a procedure for clarifier design that accounts for these interactions. This section will extend the work of Dick [2] and develop a procedure for evaluating the economic trade-offs between alternative activated sludge system designs. The authors have chosen not to employ the terminology "optimal design" because several important variables such as sludge settleability and clarification cannot be fully quantified at this time.

Design Basis/Aerator

Solids Residence Time

Design of the aeration basin of the activated sludge process should be based on effluent quality criteria. When selecting a mean solids residence time (SRT, θ_c) consideration must be given to: (a) soluble carbonaceous matter in the effluent; (b) particulate carbonaceous matter in the effluent; (c) thickening characteristics of the biological solids; (d) nutrient control alternatives; and (e) process stability.

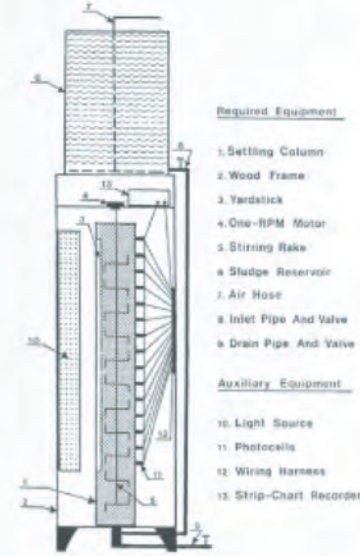

Figure 14. Schematic of test batch settling column.

If removal of carbonaceous BOD is the primary objective in selecting the design SRT value, then one must consider the variability of the first three quality criteria with respect to SRT. Figures 15, 16 and 17 are examples of such relationships. These were established by Stensel and Shell [12] and Bisogni and Lawrence [13], respectively. Only the relationship which describes the soluble carbonaceous matter in the effluent under steady-state conditions, given in Figure 15, is generally applicable for domestic wastewater. The relationships which describe thickening and clarification characteristics, given in Figures 16 and 17, are not generally applicable. These are system-specific and should therefore be established for each system that is to be investigated.

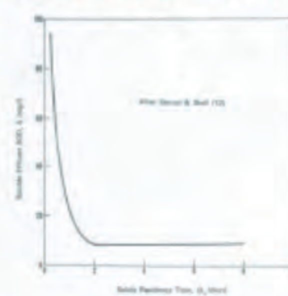

Figure 15. Soluble effluent BOD versus solids residence time.

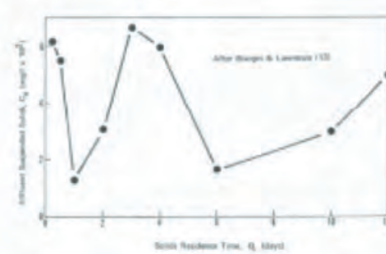

Figure 16. Effluent suspended solids versus solids residence time.

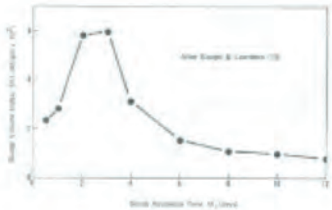

Figure 17. Sludge volume index versus solids residence time.

Upon having selected an appropriate value for solids residence time, the biooxidation model of Lawrence and McCarty [14], which defines the mean SRT in terms of reactor design parameters, (θ, C_o), influent conditions (S_o), and microbial characteristics (Y, k_d),

$$C_o = \frac{Y(S_o - S)}{1 + k_d\theta_c} \quad \frac{\theta_c}{\theta} \tag{4}$$

can be employed for establishing a design value for the aeration basin MLSS, C_o. It is to be noted that C_o varies inversely with values assumed or selected for the hydraulic residence time, θ: *i.e.,* for greater assumed design values of θ, the required MLSS, C_o, decreases proportionately. Equation 4 applies only for the case of a completely mixed aeration chamber. If a system is to be designed for a plug-flow hydraulic regime, then the design equation must be modified accordingly. Lawrence and McCarty [14] should be consulted in this regard.

Hydraulic Residence Time

Although the foregoing consideration showed that a broad range of hydraulic residence times can be employed while meeting the solids residence time design criterion, certain constraints must, on occasion, be imposed. Factors such as (a) process resilience to the imposition of shock loads, (b) kinetics of degradation of large polymeric molecules, (c) maximum aeration capacity per unit volume, (d) maximum agitation intensity per unit volume, (e) process stability, and (f) concentration of mobile ciliates must be considered when establishing constraints for hydraulic residence time. The first five must be considered when establishing a lower-level constraint on the hydraulic residence time while the last must be considered when establishing an upper limit on θ.

It is often necessary to establish a minimum hydraulic residence time requirement when designing an activated sludge system for treating industrial wastewaters. This is usually done for the purpose of accommodating shock loading inputs and providing for the degradation of large polymeric molecules. Usually, however, these two criteria are not as important when designing municipal wastewater treatment facilities. Nonetheless, the criteria regarding aeration capacity and agitation intensity must be considered in the design of municipal systems, because the quantity of air that must be supplied per unit of aeration chamber volume is larger for smaller design hydraulic residence times. The constraint regarding aeration capacity is simply a physical constraint with respect to the maximum number of diffusers of surface aerators that can be installed per unit volume of aeration chamber. Conversely, the constraint regarding agitation intensity is a performance constraint. That is, for smaller hydraulic residence times the agitation intensity in the aeration chamber is greater. Since it has been shown that excessive shear rates in the aerator tend to break up the biological floc, resulting in a degraded-quality effluent, a lower limit on the hydraulic residence time must be established to ensure that a high-quality effluent is attained.

The last factor that must be considered in establishing feasible hydraulic residence times is the concentration of mobile ciliates contained in the biosolids. Pitman [15] determined that the quality of the effluent of an activated sludge plant is materially affected by the concentration of mobile ciliates and showed that the best-quality effluent can be attained when the concentration of mobile ciliates is low and the concentration

of crawling and attached ciliates is high. Since Pitman's studies established that these conditions are favored by long solids residence times and short hydraulic residence times, consideration should be given to the population distribution of ciliates when establishing constraints on both θ_c and θ.

Design Basis/Clarifier

Five alternative equations can be employed for sizing the secondary clarifier of an activated sludge process [9]. These are listed in Table I. The first of the equations listed is the simplest and, therefore, the most convenient to use. It coordinates well with the state point concept developed earlier.

Table I. Clarifier Design Equations

Equation	Equation No.
$A = (C_oQ_i)G_o$	(5)
$A = (C_oR)/(G_L - G_o)$	(6)
$A = C_o(Q_i + R)/G_L$	(7)
$A = C_o(Q_i + R)/(C_u \cdot u)$	(8)
$A = R/u$	(9)

Once a design hydraulic residence time has been selected, thereby fixing an operational MLSS value, the clarifier can be sized by assuming a value for G_o and solving the equation for the required surface area. It is to be noted that one establishes a design state point for the system by assuming a G_o. This has the coordinates (C_o, G_o). One can then draw a recycle rate operating line which passes through the state point and is tangent to the settling flux curve as shown in Figure 12 (critically loaded case). The required minimum recycle flow rate is then calculated as the product of the clarifier surface area, A, and the recycle rate bulk transport velocity term, u.

One must recognize that only those values assumed for G_o that lie beneath the envelope of the settling flux curve at the specified C_o (MLSS) will result in technically feasible designs for the clarifier. The greater the value assumed for G_o, the smaller is the required size of the clarifier, A, and the larger is the minimum required recycle flow rate, R. It is apparent, therefore, that an economic trade-off exists between clarifier size and recycle pumping capacity for different assumed values of G_o. Values assumed for G_o are, however, subject to several constraints as discussed below.

Clarification Constraint

Design of secondary clarifiers must be based on clarification as well as thickening criteria. A clarification constraint is established by arbitrarily selecting a peak or maximum overflow rate Q_i/A, that is consistent with satisfactory operating experience. Even though the overflow rate is actually equivalent to $(Q_i - W)/A$, it is approximately equal to Q_i/A, as noted above.

Figure 18 shows five clarification or overflow rate constraints ranging from 600 to 1400 gal/ft^2/day. Technically feasible values that one could assume for G_o, therefore, can only lie within the envelope described at the upper bound by the clarification constraint and the settling flux curve.

Recent studies by Pflanz [16] and Agnew [17], were conducted to define clarification efficiencies in the activated sludge system. An analysis of the data presented by Pflanz yielded the following relation with respect to overflow suspended solids:

$$C_e = 4.5 + 1.27 \times 10^{-3} (C_o \times Q_i/A)$$

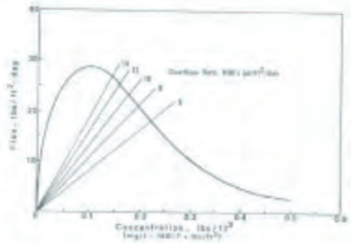

Figure 18. Overflow rate design constraints.

An analogous correlation, presented by Agnew, is of the following form:

$$C_e = 18.2 + 0.0136(Q_i/A) - 0.0033\ C_o$$

Both of these correlations show suspended solids in the clarifier effluent to be a function of the overflow rate as well as the mixed liquor suspended solids concentration, C_o. It is interesting to note, however, that while the Pflanz expression predicts higher effluent suspended solids levels, C_e, for higher mixed liquor suspended solids concentrations, C_o, the Agnew function predicts the precise converse, *i.e.*, decreasing effluent suspended solids levels for increasing C_o's. This anomaly in experimental observations cannot be resolved at this time due to the lack of an adequate data base. Nevertheless, the two functions can serve as useful empirical alternatives to the clarification constraint described above.

Figures 19 and 20 show clarification constraints developed for various desired effluent suspended solids levels, C_e, using the Pflanz [16] and Agnew [17] expressions, respectively. The constraints developed using the Pflanz function appear as horizontal lines. This is because the Pflanz constraint is, in reality, a constraint on solids surface feed, G_o, for a specified effluent suspended solids level.

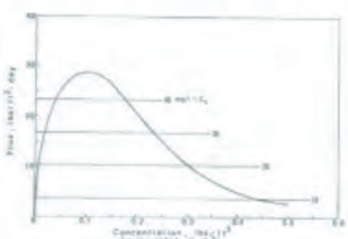

Figure 19. Overflow rate design constraints after Pflanz [16].

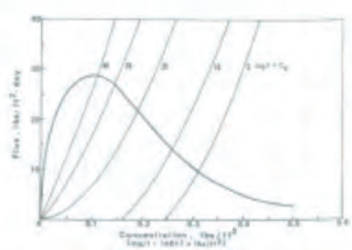

Figure 20. Overflow rate design constraints after Agnew [17].

Clarifier Underflow Concentration Constraint

It is often desirable to place a lower bound on the concentration of sludge solids that are wasted from the underflow of the secondary clarifier. In the case of the conventional activated sludge system this is usually done when no special provision is to be made for independent sludge thickening prior to digestion or some other form of sludge processing. For extended-air activated sludge treatment systems where the waste sludge is commonly discharged directly to sand drying beds, it is also desirable to place a lower constraint on the concentration of solids in the underflow of the clarifier.

Such a constraint is established by constructing a linear trace which is tangent to the settling flux curve intersecting the x-axis at the minimum desired sludge solids concentration. Three examples of such constraints are shown in Figure 21 for the cases in which

the minimum clarifier underflow suspended solids concentration is limited to 6407, 8009 and 9611 mg/l (0.4, 0.5 and 0.6 lb/ft^3, respectively).

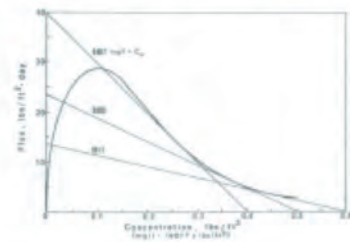

Figure 21. Underflow concentration design constraints.

The underflow concentration constraint is an upper bound for design state points (C_o, G_o) which lie to the left of the point at which the constraint line and the settling flux curve are tangent. For state points that lie to the right of the point of tangency, the settling flux curve itself is the upper boundary, because the underflow concentration constraint is automatically satisfied in such cases.

Recycle Rate Constraint

When evaluating economic trade-offs between alternative activated sludge system designs one occasionally arrives at an apparent least-cost design for which the required recycle rate may not be realistic. That is, required recycle flow rates for the lowest-cost design case can be as high as 300 to 400% of the influent flow rate. Although the cost calculations include the capital costs associated with the installation of recycle pumping capacity and the operating costs of recycle pumping,no consideration is given to the design procedure relative to the effect high recycle pumping rates have on the quality of the effluent. While no direct evidence is available, high recycle pumping rates probably degrade the quality of the effluent by increasing turbulence and therefore decreasing clarification efficiencies in the secondary clarifier. Consequently, it may be desirable to impose recycle pumping rate constraints on the feasible design domain.

Two recycle flow rate constraints, 50 and 100% of the influent flow rate, are shown in Figure 22. These were constructed by evaluating the following inequality,

$$\frac{u}{G_o} \leq \frac{(R/Q_i)}{C_o}$$

employing a trial and error procedure. This constraint, as well as the underflow concentration constraint, is a function of the settling flux curve. If a recycle flow rate constraint is to be imposed, it is an upper bound for the feasible design domain.

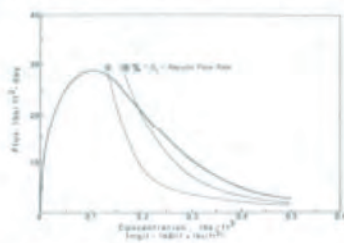

Figure 22. Recycle flow rate design constraints.

Because considerable effort is required to develop a recycle rate constraint, it is desirable to first evaluate the design alternatives without the recycle rate constraint. If the resulting lowest-cost design requires an unacceptably high recycle pumping rate, the design should be repeated with a recycle rate constraint included.

Application of Clarifier Constraints

A clarification constraint must be considered in each design. If these were the only constraints applied, the feasible design domain would consist of the shaded area shown in Figure 23 (Q_i/A = 800 gal/ft^2/day). The alternative design operational points that are described by the coordinates (C_o, G_o) must lie within the shaded region. All alternative designs resulting from assumed operational points that lie outside the shaded area would either violate the clarification criterion or fail due to insufficient thickening capacity.

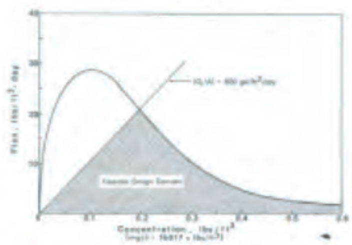

Figure 23. Feasible design domain with clarification constraints only.

If analysis of the feasible alternative designs showed that the least-cost solution required a recycle flow rate that was unacceptably high, one would superimpose a recycle rate constraint as shown in Figure 24 for the case in which the recycle rate was limited to 100% of the influent flow rate. The feasible design domain is again shown by the shaded region.

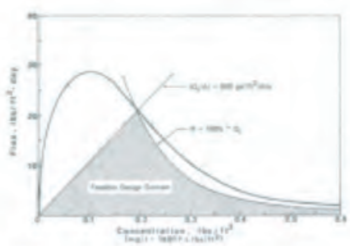

Figure 24. Feasible design domain with clarification and recycle rate constraints.

If, in addition to a clarification constraint, certain process considerations dictated the imposition of an underflow concentration constraint at 0.5 lb/ft^3 (8009 mg/l), feasible design domain would consist of the shaded area shown in Figure 25. As before, if it were

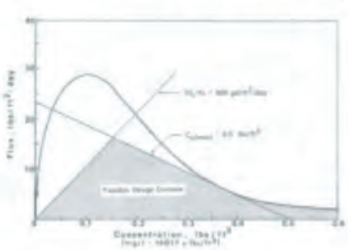

Figure 25. Feasible design domain with clarification and underflow concentration constraints.

also necessary to impose a recycle rate constraint, the feasible design domain would decrease in size to that shown in Figure 25.

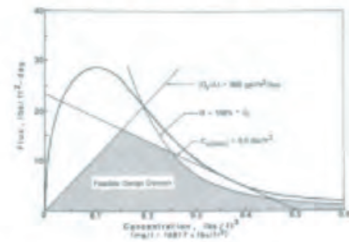

Figure 26. Feasible design domain with clarification, recycle rate, and underflow concentration constraints.

Illustrative Design Example

The design methodology presented above can best be illustrated by use of a comprehensive design example. For this example it has been assumed that a completely mixed activated sludge treatment plant is to be designed for an average flow of 10.0 mgd and a BOD_u of 300 mg/l in the primary clarified wastewater. Based on an analysis of the functional relationships shown in Figures 16, 17 and 18, a mean solids residence time of 6 days is to be employed for the design. All cost calculations were based on cost relationships established by Patterson and Banker [18]. Other salient design parameters are tabulated in Table II.

Table II. Design Parameter Values

Influent Parameters	Physical Parameters
Q_i = 10.0 mgd (average influent flow rate)	hydraulic regime: complete-mix
S_o = 300 mg/l BOD_u	oxygen transfer efficiency: 8%
Effluent Parameters	Cost Parameters
$S \leq$ 8 mg/l BOD_u (soluble)	interest rate = 6% per annum
Biologic Parameters	depreciation period = 25 yr
θ_c = 6 days (design mean solids residence time)	power costs = 1.48¢/kWh
Y = 0.5 mg VSS/mg BOD_u (true yield coefficient)	operations labor costs[a] = \$6.92/man-hr
k_d = 0.06/day (biomass decay rate)	maintenance labor costs[a] = \$6.89/man-hr
MLVSS = 0.8 MLSS	cost index[b] = 1 January 1976 dollars
Constraints	
Q_i/A = 800 gal/ft^2/day (average flow basis; provides for peak flows)	
aerator mixing: minimum of 3 cfm per foot of chamber length	
aeration capacity: install 150% of minimum required capacity	
recycle pumping capacity: install 150% of minimum required capacity.	

[a]Ref: Joseph W. Larson, Superintendent of Coldwater Creek Wastewater Treatment Plant, Metropolitan Sanitary Treatment Plant, Metropolitan Sanitary District, Annual Report 1975-76, St. Louis, Missouri.

[b]Means Construction Cost Indexes, 2, 1 (1976).

Assuming that the expected settling flux curve for the biological sludge is as shown in Figure 27 and assuming an overflow rate constraint of 800 gal/ft^2/day, the feasible design domain is described by the shaded region. It is assumed further that no underflow concentration constraint need be imposed in this case.

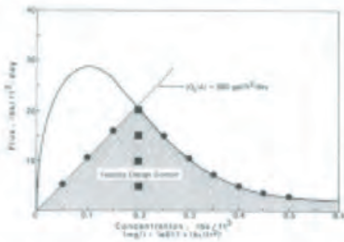

Figure 27. Feasible design domain for example design problem.

As indicated above, a spectrum of clarifier size/recycle flow rate design combinations is feasible at each specified hydraulic residence time (*i.e.*, specified concentration of mixed liquor suspended solids). The relative magnitude of the values determined for clarifier size and recycle flow rate depend on the value assumed for G_o: larger G_o's yield smaller values for required clarifier size and larger values for required recycle flow rate. Since this is the case, one must establish which of the combinations is the least-cost combination for each specified hydraulic residence time.

To establish the trade-off cost relationships for the critical case of θ equal to 6 hr ($C_o = 0.2$ lb/ft^3, 3203 mg/l), cost comparison calculations were conducted for five selected values of G_o. The design operating points with coordinates (C_o, G_o) are shown in Figure 27 as solid squares while the associated costs for the clarifier/recycle system are recorded in Table III. Analysis of the total annual costs shows that the cost decreased

Table III. Least-Cost Design Analysis For a Specific Hydraulic Time

	G_o(lb/ft^2/day)				
Variable	5	10	15	20	20.7
		Design Values			
A (ft^2)	53,475	26,738	17,825	13,369	12,916
u (ft/day)	8.4	26.4	52.5	98.7	115.3
R (mgd)	3.36	5.28	7.00	9.86	11.13
		Clarifier Annual Costs ($)			
Capital Amortization	62,848	35,497	25,023	20,367	19,785
Operating Labor	19,376	12,802	9,688	7,958	7,612
Maintenance Labor	11,024	6,890	5,236	4,478	4,134
Material	8,480	4,909	3,571	2,827	2,752
TOTAL	101,728	60,098	43,518	35,630	34,283
		Recycle Pumping Annual Costs($)			
Capital Amortization	8,496	11,406	13,966	17,458	18,221
Operating Labor	3,667	4,083	4,360	4,844	5,052
Maintenance Labor	3,376	3,513	3,652	3,996	4,203
Power	2,976	4,463	5,504	7,587	7,736
Other	922	1,175	1,710	2,529	2,540
TOTAL	19,437	24,640	29,192	36,414	37,752
		Total Annual Costs			
Dollars	121,165	84,738	72,710	72,044	72,035
Cents/1000 gal	3.32	2.32	1.99	1.97	1.97

for larger selected values of G_o up to 20.7 lb/ft^2/day, although the cost relationship is insensitive for G_o values ranging from 15 to 20.7 lb/ft^2/day. For this hydraulic residence time (θ = 6 hr), therefore, the least-cost solution was obtained for the case in which the design state point was located at the upper boundary of the feasible design domain. Moreover, for all other aerator hydraulic residence time alternatives evaluated for this example, the least-cost solution occurred at the upper boundary of the design domain.

Nevertheless, it is extremely important that one recognizes that the least-cost solution does not *a priori* lie at the upper boundary of the design domain; it can lie anywhere within the domain. At a certain θ the location of the design state point for the least-cost solution for the clarifier/recycle pumping system depends on the specifics of the system design and the unit cost relationships.

Annual costs calculated for the aeration basin/clarifier system for assumed hydraulic residence times ranging from 2.41 to 24.1 hr are recorded in Table IV. At each hydraulic residence time for which costs were calculated the least-cost design state points are shown as solid circles in Figure 27. These are located at the upper bound as established earlier.

Analysis of the annual costs calculated for the aeration basin and aerator and for the clarifier and associated recycle pumping, plotted in Figure 28, shows that the least-cost solution was obtained for a hydraulic residence time of approximately 8 hr. One can observe that the total cost function is extremely sensitive for low hydraulic residence times and is considearbly less sensitive for higher hydraulic residence times. This is primarily due to the clarifier and recycle pumping cost function.

The required recycle flow rate was 44.1% of the influent flow for the least-cost solution. Since this is commensurate with operations practices that are employed at treatment plants producing high-quality effluents, it was not necessary to apply a more stringent recycle rate constraint. Nonetheless, had the least-cost solution required an unrealistically high recycle flow rate, then a recycle flow constraint would have had to have been applied. This would have increased the total annual costs for small hydraulic residence times, thereby shifting the least-cost solution toward higher θ's.

No underflow concentration constraint was applied for this example. If one had been applied, however, it also would have served to increase the total annual costs for small hydraulic residence times and would have shifted the least-cost solution toward higher θ's.

It is noteworthy, furthermore, that one could extend the methodology presented herein to evaluate the economic alternatives for the entire aeration basin/clarifier/sludge processing system. This would be achieved by including sludge processing costs in the least-cost calculations and by iteratively applying a spectrum of feasible underflow concentration constraints.

OPERATIONAL STATE ANALYSIS AND CONTROL

As indicated in the introductory paragraphs, the batch settling approach can be used by operations personnel to monitor the operational state of an activated sludge system. Biosolids inventory control decisions can subsequently be made in response to both short-term (diurnal) and long-term (seasonal) changes in the plant influent and system operational conditions. The approach is to be employed in conjunction with a suitable SRT solids wasting control strategy such as that proposed by Garrett [19] for direct wasting from the aerator or by Zaander and Johnson [20] and Walker [21] for systems in which sludge is wasted from the settler underflow. A more precise solids wasting strategy that includes the loss of suspended solids in the plant effluent, as presented by Roper and Grady [22], could also be employed. Any of these can be used to effectively control an activated sludge system at a desired set-point solids residence time that provides the desired performance and process stability.

Within the constraint of a desired solids residence time which establishes the sludge wasting program, the treatment plant operator can control only the recycle flow rate to control the biosolids inventory in the aerator/clarifier system of a conventional activated sludge treatment plant. The objectives of recycle rate control are: (a) to prevent clarifier failure due to overload, and (b) to produce a thickened sludge to minimize sludge processing and disposal costs. Within these objectives, moreover, it is desirable to minimize

Table IV. Least-Cost Design Analysis

Variable	Hydraulic Residence Time (θ, hr) 24.12	12.06	8.04	6.03	4.82	4.02	3.45	3.02	2.68	2.41
				Design Values						
C_0 (#/ft^3)	0.05	0.10	0.15	0.20	0.25	0.30	0.35	0.40	0.45	0.50
C_0 (mg/l)	801	1,602	2,403	3,203	4,005	4,806	5,607	6,407	7,208	8,009
A (ft^2)	12,500	12,500	12,500	12,916	22,281	38,938	64,988	115,779	171,886	267,380
u (ft/day)	6.37	20.09	47.2	115.26	100.0	78.52	56.6	39.8	20.09	6.37
R (mgd)	0.595	1.88	4.41	11.1	16.7	22.9	27.5	33.9	25.8	12.7
V (Mgal)	10.05	5.02	3.35	2.51	2.01	1.67	1.44	1.26	1.12	1.01
Air Required (cfm)	8957	5808	5808	5808	5808	5808	5808	5808	5808	5808
				Aeration Basin and Aerator Annual Costs						
Basin Capital Amortization	151,291	67,509	48,266	36,063	30,274	26,206	22,138	19,791	18,618	17,445
Aerator Capital Amortization	33,716	33,168	33,168	33,168	33,168	33,168	33,168	33,168	33,168	33,168
Operating Labor	17,992	17,992	17,992	17,992	17,992	17,992	17,992	17,992	17,992	17,992
Maintenance Labor	10,335	10,335	10,335	10,335	10,335	10,335	10,335	10,335	10,335	10,335
Power	52,073	52,073	52,073	52,073	52,073	52,073	52,073	52,073	52,073	52,073
Other	8,778	8,778	8,778	8,778	8,778	8,778	8,778	8,778	8,778	8,778
TOTAL	274,185	189,855	170,612	158,409	152,620	148,552	144,484	142,137	140,964	139,791
				Clarifier and Recycle Pumping Annual Costs						
Clarifier Capital Amortization	19,786	19,786	19,786	19,786	30,260	48,882	75,651	110,567	174,579	256,049
Pumping Capital Amortization	3,607	6,052	10,241	18,221	26,769	34,920	40,735	47,136	37,243	22,113
Clarifier Operating Labor	7,266	7,266	7,266	7,612	12,110	15,916	22,144	31,140	40,136	50,516
Clarifier Maintenance Labor	4,065	4,065	4,065	4,134	6,201	8,612	12,402	17,914	22,048	27,560
Pumping Operating Labor	3,128	3,391	3,944	5,052	6,020	7,612	8,304	8,966	7,612	5,536
Pumping Maintenance Labor	2,549	2,825	3,307	4,203	4,961	5,994	6,545	7,579	6,270	4,616
Clarifier Material	2,700	2,700	2,700	2,752	4,315	6,695	9,819	14,878	20,829	28,268
Pumping Power	1,042	2,083	4,017	7,736	11,902	16,366	18,597	25,293	17,100	9,075
Pumping Other	461	640	1,175	2,540	3,868	5,505	6,100	8,183	5,700	2,926
TOTAL	44,604	48,808	56,501	72,035	106,406	150,502	200,297	271,686	331,517	406,709
				Total Annual Costs						
Dollars	318,789	238,663	227,113	230,444	259,026	299,054	344,781	413,823	472,481	546,500
Cents/1000 gal	8.73	6.54	6.22	6.33	7.09	8.19	9.44	11.44	12.94	14.97

sludge processing and disposal costs. Within these objectives, moreover, it is desirable to minimize the recycle flow rate in order to minimize recycle pumping operating costs. This strategy should also result in enhanced clarification efficiencies since small recycle flow rates would create less turbulence in the clarifier.

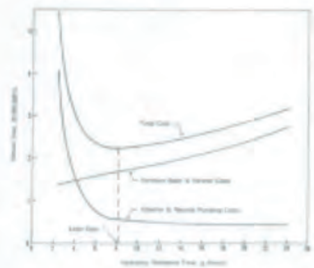

Figure 28. Costs for activated sludge treatment system.

Analysis of Equation 4 shows that the concentration of mixed liquor suspended solids, C_0, varies as a function of the concentration of substrate in the influent, S_0, and the hydraulic residence time, θ, for a constant θ_c. Such variations in C_0 due to organic loading can easily be included in analyzing the operational state of a system, although the analytical procedure is slightly complicated. However, it is usually not necessary to consider these variations when analyzing the average changes in the state of a system, provided the mean daily value of C_0 does not change appreciably.

To illustrate how a treatment plant operator would establish the operational state of a system and make control decisions, three different types of process changes or disturbances are considered. These are (a) changes in the influent flow rate, (b) changes in solids residence time, and (c) changes in the settling characteristics of the biological sludge. The analysis and control procedure can best be illustrated by a comprehensive example. For this purpose it has been assumed that the plant that was designed in the preceding section (least-cost basis) is to be operated. It is recalled that the plant was designed for an average hydraulic residence time of 8.04 hr and a mean solids residence time of 6 days. The final design called for an aerator volume of 3.35 million gal (completely mixed hydraulic regime) and an installed clarification capacity of 12,500 ft^2 of surface area. It has been assumed, furthermore, that the current operating conditions are:

1. Average daily influent flow rate = 10.29 mgd
2. Maximum daily influent flow rate = 12.86 mgd
3. Minimum daily influent flow rate = 7.72 mgd
4. Mean solids residenct time (θ_c) = 5.77 days
5. Recycle flow rate (R) = 4.39 mgd (42.7% of average daily influent flow)
6. MLSS (C_0) = 2403 mg/l (0.15 lb/ft^3)

The assumed operating conditions are noted to be somewhat different than those employed for design. This was done deliberately to enable illustration of the utility and versatility of the approach.

Variable Influent Flow Rate

Since essentially all wastewater treatment plants experience both short-term (diurnal), intermediate-term (seasonal), and long-term (design period) changes in the influent flow rate, it is instructive to consider how these changes affect the operational state of the plant and how a plant operator should respond to the flow changes that are imposed. For this example the base operational state is established for the case in which the plant is experiencing an average daily hydraulic rate of flow, 10.29 mgd. For this case the operational point is located as shown in Figure 27 and is described by state point (a) which has the coordinates (0.15 lb/ft^3, 16.5 lb/ft^2/day). The recycle rate operating line has a

slope equal to 47 ft/day (labeled u_1) while the overflow rate operating line has a slope equal to 823 gal/ft^2/day (labeled Q/A_{avg}). Because the recycle rate operating line is tangent to the settling flux curve, the clarifier is critically loaded.

Increased Hydraulic Flow Rates

Although influent flow rates seldom change stepwise, the flow rate has been assumed to change from the average daily flow, 10.29 mgd, to the maximum daily flow, 12.86 mgd, in a single step for illustrative purposes. The state point is seen to shift upward in Figure 29 at the same C_0 (MLSS) to state point (b) which has coordinates of (0.15 lb/ft^3, 20.62 lb/ft^2/day). Due to the increased flow the overflow rate operating line has an increased slope, 1028 gal/ft^2/day. As before, the wasting flow rate, w, is neglected when determining clarifier overflow rates. Since the recycle rate was not changed, the new recycle rate operating line (dashed) has simply shifted upward and has the same slope as before the step change in flow was imposed.

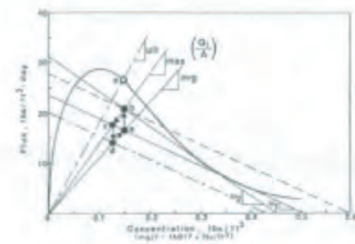

Figure 29. Operational states for increased influent hydraulic flow rates.

Analysis of the new operating state shows that the clarifier is overloaded with respect to its thickening capacity as the recycle rate operating line passes above the settling flux curve. If this condition were allowed to persist, biological solids would be transferred from the aerator and would accumulate in the clarifier. To resolve this condition, the operator could increase the recycle flow rate such that the recycle flow rate operating line has a slope equal to 69.33 ft/day (u_2). This corresponds to a recycle flow rate of 6.48 mgd or 50.4% of the influent flow. The resulting operating state corresponds to a critically loaded state for the clarifier. This is a stable state for which the required recycle flow rate is at a minimum.

Conversely, if the plant operator took no control action in response to the imposed maximum daily hydraulic flow, then the clarifier's thickening capacity would be overloaded as mentioned above. The system would itself respond by transferring biological solids from the aerator to the clarifier, thereby diluting out the solids in the aerator resulting in a decreased concentration of mixed liquor suspended solids. The solids "blanket" in the clarifier would rise to accommodate the solids that were transferred.

This shift in operational states is shown in Figure 29. From the overloaded state (state point b) the state point would shift down along the overflow rate operating line until it reached the location of point (c), which corresponds to a stable operational state for the aerator/clarifier system for which the recycle rate was not changed. Thereafter, no further change in the operating state of the system would be observed; *i.e.*, no additional biosolids would be transferred from the aerator to the clarifier.

To calculate the mass of solids transferred during the shift from the overloaded state to the critically loaded state (point b to point c), one would simply graphically establish the original and final concentrations of mixed liquor suspended solids, C_0, take the difference and multiply by the aerator volume. For this case the original and final C_0 values were 2403 mg/l (0.15 lb/ft^3) and 2082 mg/l (0.13 lb/ft^3), respectively. One can then calculate that approximately 8957 lb of biosolids would be transferred from the aerator to the clarifier.

It is also interesting to establish precisely how the clarifier would accommodate the solids. As noted previously, the solids would accumulate in the clarifier in the form of a "blanket" at a concentration equal to C_L which has a value of 6087 mg/l (0.38 lb/ft^3) for the current operating conditions. Since 4750 lb of biosolids (0.38 lb/ft^3 x 12,500 ft^3) can be stored in the clarifier per foot of clarifier depth at that concentration, it is apparent that the solids blanket would rise approximately 1.9 ft in the clarifier and then stabilize. Because clarifiers usually have 6 to 8 ft of depth available for storage of biosolids beneath the feed point, one notes that the clarifier would not fail for the transition in operational states considered.

If the influent flow rate to the plant subsequently changed stepwise from the maximum value, 12.86 mgd, back to the average value, 10.29 mgd, the operational point would correspondingly shift in Figure 29 from state point (c) to state point (d) which has the coordinates (0.13 lb/ft^3, 14.3 lb/ft^2/day). This operating condition is one in which the clarifier is underloaded, as evidenced by the recycle rate operating line (long and short dashes) which passes beneath the settling flux curve. Solids that are stored in clarifier in the "blanket" consequently would be transferred from the clarifier back into the aerator. This would result in a shift of operational states from state point (d) to state point (a). At the latter state point the blanket in the clarifier would have been entirely dissipated.

Clearly, clarifiers offer considerable capacitance for accommodating the usual diurnal fluctuations in influent hydraulic flow rate. Only for certain critically located state points that are near the settling flux curve would one expect the blanket to propogate upward in the clarifier to the extent that solids would be scoured from the blanket over the weirs and into the effluent.

Decreased Hydraulic Flow Rates

If the flow to the treatment system were to decrease, then the state point would shift downward at the same C_0 (MLSS). This is illustrated in Figure 30 for the case in which the influent flow rate changed stepwise from the average, 10.29 mgd, to the daily minimum flow of 7.72 mgd (state point (a) to state point (b)). If the treatment plant operator did nothing in response to the change in flow, the recycle rate operating line would

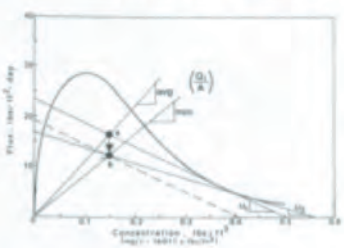

Figure 30. Operational states for decreased influent hydraulic flow rates.

be positioned as shown by the dashed line. It is noted, however, that for this operational state the clarifier is underloaded. The recycle rate could accordingly be reduced to the point of critical loading: *i.e.*, decrease the slope of the recycle rate operating line to 30.1 ft/day (u_3) which corresponds to a recycle flow rate of 2.82 mgd or 36.5% of the influent flow. This control action is beneficial as it results in reduced recycle pumping operating costs and reduced turbulence in the clarifier which should, in turn, result in upgraded clarification efficiencies.

For all situations in which the clarifier is underloaded the vast majority (95+%) of the biosolids are in the aerator and no redistribution of solids between the aerator and clarifier occurs. Accordingly, the concentration of MLSS should vary only in response to changes in organic loading to the aerator, provided, of course, that a carefully controlled solids wasting procedure is employed to maintain a constant set-point solids residence time.

Flow Proportional Control

Many wastewater treatment plants are operated with flow-proportional control of the recycle pumps. If this control strategy had been employed for the example discussed above, then the recycle flow rates that would have been used based on a 42.7% recycle ratio are 3.30, 4.39 and 5.49 mgd, respectively. Comparing these with the recycle flow rates required to maintain the clarifier in a critically loaded condition one observes that for the minimum influent flow the flow-proportional recycle rate was somewhat greater than required (Table V). Conversely, the proportional recycle flow was smaller than the required recycle flow for the maximum influent flow case. For the average flow case the two compared precisely since the basis of selecting the ratio was the case of average flow and critical loading. It is apparent, therefore, that while the flow-proportional recycle control strategy did not provide for the precise recycle rate required to maintain the clarifier in a critically loaded condition, it did provide a relatively good estimate of the required flow. Accordingly, one can conclude that this control strategy could be employed as an alternative approach for minimizing set-point recycle flows. It would provide for relatively stable diurnal operation of an aerator/clarifier system if the recycle ratio were properly selected.

Table V. Recycle Rate Comparison for Three Influent Flow Cases

Influent Flow Case	Q_1/A (gpd/ft^2)	Required Recycle Rate (mgd/%)	Flow-Proportional Recycle Rate (mgd/%)
Minimum (7.72 mgd)	617	2.82/36.5	3.30/42.7
Average (10.29 mgd)	823	4.39/42.7	4.39/42.7
Maximum (12.86 mgd)	1028	6.48/50.4	5.49/42.7

If the flow-proportional recycle ratio had been selected at some value greater than that determined for the average flow/critical loading case, then a different control situation would result. To illustrate, it is assumed that a recycle ratio of 50% is to be employed. Figure 31 shows the three operating states for the minimum, average and maximum influent flow rate cases described above. In each case the recycle flow rate is greater than

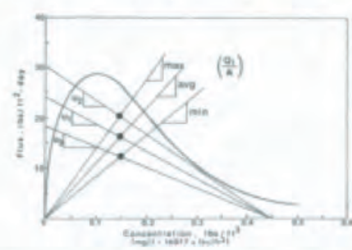

Figure 31. Operational states for flow-proportional recycle rate control.

required and in no case is the clarifier overloaded or even critically loaded. It is to be noted, furthermore, that the clarifier underflow or recycle line solids concentration remains unchanged regardless of the operational state of the system. This recycle control strategy would provide for stable diurnal operation of an aerator/clarifier system. Nonetheless, the operational costs for this strategy would be slightly higher than for the control strategy described above. Also, a slightly greater degree of turbulence would be imposed on the clarifier.

Hydraulic Washout

Another factor that must be considered relates to the state point, shown in Figure 29, that lies at the intersection of the settling flux curve and the overflow rate operating line labeled $(Q/A)/_{ult}$. At this state point the initial interface settling velocity (ISV) of the sludge at the MLSS concentration is equal to the overflow rate, 180 ft/day (1346 gal/ft^2/day). For all higher hydraulic loadings, consequently, solids introduced to the clarifier would not settle into the sludge zone but would be propagated upward from the feed point and over the weirs. This is commonly referred to as "hydraulic washout."

To illustrate this phenomenon it is instructive to consider an example. As for the previous examples, the base case is developed for the situation in which the plant is experiencing an average daily hydraulic flow of 10.29 mgd. For this case the state point (labeled point a) is located as shown in Figure 32. Due to the occurrence of an extreme rainfall it is assumed that the flow rate to the plant surged rapidly to 23.99 mgd. At the new state point (point b) it is apparent that the plant operator has no control alternatives available to him other than by-passing the secondary system. No recycle rate control options could be employed to resolve this hydraulic overload problem.

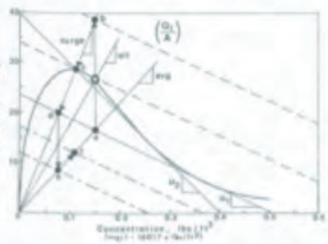

Figure 32. Operational states for hydraulic washout case.

Since the biosolids that are introduced to the secondary clarifier settle more slowly than the imposed hydraulic overflow rate, they are transported from the feed point upward into the effluent. This results in a shift to the state point from point (b) downward along the (Q_i/A) surge overflow rate operating line as more and more biosolids are discharged over the weirs. When the state point reached location (c), coordinates of (0.115 lb/ft^3, 28.7 lb/ft^2/day), a total of 15,675 lb of biosolids would have been discharged over the weirs. Because the hydraulic overflow rate and the interface settling rate of the diluted biosolids in the aerator are equal at state point (c), no additional solids would inadvertently be carried over the weirs.

The plant operator again has recycle rate control options available to him at state point (c). Assuming that the operator chose not to change the recycle rate, however, then the system which is overloaded with respect to thickening capacity would itself respond by transferring solids from the aerator to the clarifier, thereby further diluting out the solids in the aerator. This shift in operational states is shown in Figure 32 as a transition from state point (c) to point (d) (0.078 lb/ft^3, 19.8 lb/ft^2/day). The latter operating condition corresponds to a stable operational state for the aerator/clarifier system for which the recycle rate had not been changed. During the transition from state point (c) to point (d) a total of 16,571 lb of biological sludge would have been transferred from the aerator to the clarifier. This would have resulted in a 3.5-ft rise of the "blanket" in the clarifier. At state point (d) the concentration of mixed liquor suspended solids would have dropped to 1249 mg/l from the original level of 2403 mg/l. Of the total of 32,246 lb of solids transferred from the aerator 15,675 lb would have been permanently lost over the weirs.

If the influent hydraulic flow rate then decreased to the average flow rate, 10.29 mgd, after the system had attained a steady-state condition at state point (d), then the state point would shift to point (e). At this state point, however, the clarifier is underloaded and the biosolids stored in the "blanket" of the clarifier would consequently be transferred from the clarifier to the aerator. This would result in a shift to the state point to

point (f) in Figure 32. At the latter state point the blanket in the clarifier would have been entirely dissipated and C_o would have increased to 1842 mg/l.

Conversely, if the plant operator had elected to exercise a recycle rate control option when the state point had reached point (c), he would have increased the recycle flow rate to 9.23 mgd (38.5% of the increased influent flow) as shown by the recycle rate operating line which has a slope of u_2 (98.7 ft/day). This is a stable operational state for which the required recycle flow rate is at a minimum. The transfer of solids from the aerator to the clarifier would then have been prevented. Nothing, however, could have been done to prevent the discharge of solids over the weirs.

In addition to using the foregoing analysis of "hydraulic washout" for operations purposes, the analytical approach could also be employed from a design viewpoint. That is, one could establish the magnitude of a hydraulic surge that could be accommodated by the system without indiscriminate solids wasting. Provision for interim storage would have to be made for all hydraulic surges exceeding the identified maximum.

Changes in Sludge Settling Characteristics

In the preceeding analyses it was assumed that the settling characteristics of the sludge did not change in the time-scale considered. It must be recognized, nevertheless, that sludge settling characteristics do change seasonally and often in much shorter time domains (*e.g.*, monthly, weekly and daily). Such changes can usually be attributed to changes in either solids residence time, temperature or biomass population distributions.

Two settling flux curves are shown in Figure 33; one for the base case and the other for a degraded settling sludge. If one assumed that the activated sludge system were operated for the base case as described by the state point having coordinates of (0.15 lb/ft^3, 16.5 lb/ft^2/day), and the two operating lines of slopes, $(Q_i/A)_{avg}$ and u_1, and that the

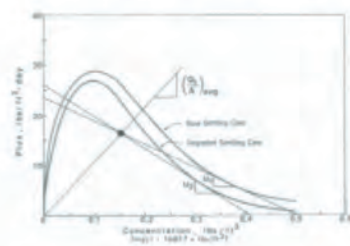

Figure 33. Operational states for changed settling characteristics.

settling characteristics subsequently deteriorated, then it is apparent that the clarifier would be overloaded. To prevent the transfer of biosolids from the aerator to the clarifier, the operator could simply increase the recycle flow rate from 4.30 to 5.98 mgd (58.1% of the influent flow) which corresponds to a u_2 of 64 ft/day and the condition of critical loading.

It is again important to underscore the statement that settling flux curves must be updated frequently enough to provide an accurate description of the current sludge settling characteristics. The required frequency for such updating measurements depends on the sludge itself as well as the physical, chemical and biological characteristics of the wastewater and the operational state of the activated sludge system. Only with truly current settling flux curves can the utility of the operational state analysis and control approach established herein be realized.

Changes in Solids Residence Times

It is occasionally desirable to change the operational set-point of solids residence time for an activated sludge system. This is easily accomplished by changing the solids wasting program. Such changes in operational state can be illustrated using the batch settling flux operations analysis approach. To do so it has been assumed, for example, that θ_c is to be changed from 5.8 days (base case) to 7.1 days. This, of course, is accomplished by decreasing the rate at which biosolids are wasted from the system, which results in an increase in the concentration of mixed liquor suspended solids. Figure 34 illustrates the transition; *i.e.*, the operating point shifts along the overflow rate operating line from state point (a) to point (b).

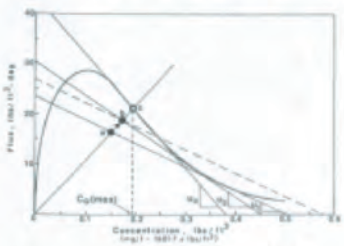

Figure 34. Operational states for increased solids residence times.

If the plant operator did not change the recycle flow rate during the transition, (dashed line) it is apparent that the clarifier would be overloaded at state point (b) and that the net increase in the mass of biosolids would be transmitted to the clarifier and stored therein. To retain those solids in the aerator, the operator would have to increase the recycle flow rate to 62.8% of the average influent flow (u_2 = 68 ft/day). This corresponds to a stable operational state. The salient required operating conditions have been tabulated in Table VI.

Table VI. Recycle Rate Comparison for Three SRT Values

Case	SRT (θ_c) (days)	MLSS (C_o) (mg/l)	Required Recycle Rate (R) (%)
Avg	5.8	2403	42.7
High	7.1	2803	61.8
Low	3.8	1762	23.6
Max	8.2	3076	104

Conversely, if the operator wanted to decrease the solids residence time set-point from 5.8 to 3.8 days, he would simply increase the solids wasting program. This would result in a net loss of biosolids and, therefore, a lower concentration of MLSS. This transition is illustrated in Figure 35 by a shift of the state point along the overflow rate operating line from state point (a) (base case) to point (d). It is apparent that the recycle flow rate at state point (d) is higher than required (dashed line). To minimize recycle pumping operating costs and the degree of turbulence imposed on the clarifier, the recycle flow rate could be decreased to 23.6% of the 10.29 mgd average influent flow rate (u_3 = 26 ft/day).

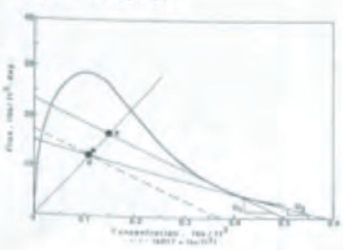

Figure 35. Operational states for decreased solids residence times.

Another consideration that must be established relates to the state point, shown in Figure 34, that lies at the intersection of the settling flux curve and the overflow rate operating line, state point (c). At this state point the initial settling velocity (ISV) of the sludge at a concentration of 3076 mg/l is equal to the average overflow rate of the clarifier; 110 ft/day (823 gal/ft^2/day). If the MLSS concentration were permitted to increase further or if the flow rate to the system increased, the biosolids would not settle into the thickening zone of the clarifier. Rather, as indicated earlier, the biosolids would be propagated from the feed point upward and would wash out over the weirs. Consequently, state point (c) establishes the maximum θ_c and the corresponding C_o at which the biological system could be operated in a stable condition. For the current example θ_c is limited to a maximum of 8.2 days. This corresponds to a C_o of 3076 mg/l at the average hydraulic flow rate.

It is to be noted, however, that a sludge's settling characteristics usually change when θ_c is changed. Accordingly, one would have to consider such changes when conducting an operational state analysis with respect to changes in the solids residence time.

SUMMARY

Use of a unified systems approach for designing and operating an activated sludge system has been illustrated. The approach enables design engineers to evaluate the economic trade-offs between alternative system designs and to establish a least-cost design. Once a system has been constructed, the same approach can be employed by plant operations personnel to establish the operational state of the system and subsequently to make rational decisions regarding required control actions or responses.

Although the approach was illustrated for the case in which the aeration basin had a completely mixed hydraulic regime, the same methodology would be employed for other hydraulic regimes in the aerator.

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NOTATION

- A = surface area of secondary clarifier, L^2.
- C_o = concentration of biosolids in aeration basin, ML^{-3}
- C_e = effluent suspended solids in clarifier overflow ML^{-3}
- C_L = solids concentration in thick blanket, ML^{-3}
- C'_L = solids concentration in dilute blanket, ML^{-3}
- C_u = solids concentration in underflow, ML^{-3}
- G = total flux, ML^{-2} T^{-1}
- G_o = solids flux due to influent flow, ML^{-2} T^{-1}
- G_b = bulk flux, ML^{-2} T^{-1}
- G_s = settling flux, ML^{-2} T^{-1}
- k_d = decay rate of biosolids, T^{-1}
- Q_i = flow rate to plant, L^3 T^{-1}
- R = recycle flow rate, L^3 T^{-1}
- S = concentration of soluble substrate in effluent, ML^{-3}
- S_o = concentration of substrate in influent, ML^{-3}
- u = underflow withdrawal velocity, LT^{-1}
- v = gravitational sedimentation velocity, LT^{-1}
- V = volume of aeration chamber, L^3
- w = waste flow rate, L^3 T^{-1}
- y = true yield coefficient, MM^{-1}
- θ = hydraulic residence time, T
- θ_c = solids residence time, T

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