Complexity versus Simplicity: A Time and a Place for All Models

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ABSTRACT
Operators have many tools for solving problems with process operations. The oldest method for problem solving relies on current operating information combined with operator training and experience. A more recent and very sophisticated approach involves the use of a calibrated process model that can be run under the current plant operating conditions. These whole-plant, commercial process models are invaluable for evaluating alternative process configurations for design and upgrade and provide an aid for assessing different control options. Models also have the potential to solve process problems and can be developed and maintained as operator training tools. There is, though, a commitment of time and financial resources required to calibrate and maintain complex process models that may be out of reach for small to mid-sized plants.

An alternative to these methods is to use a customized, but simpler, process calculator based on the plant configuration. Use of a process calculator allows for a science-based approach to assist the operator in decisions based on current and/or projected operating conditions.

This paper provides background on whole-plant commercial process models and process calculators and identifies ideal situations for using these computing tools. Software tools used at the Nine Springs Wastewater Treatment Plant (NSWTP), Madison, Wisconsin and other smaller facilities, serve as case studies.

KEYWORDS: Activated Sludge Model, Clarifier Capacity, State Point Analysis, Process Calculator, Activated Sludge Troubleshooting

INTRODUCTION AND THE POWER OF PROCESS MODELS
Whole-plant process modeling is a beneficial tool for engineering design and operations problem solving for treatment plants. Whole-plant process models can iteratively run through process calculations to find an "end-state" for a dynamically changing process. For example, influent and solids loading that is removed to a digestion process can later add a filtrate load to a plant that ultimately needs to be accommodated in treatment operations. The generation of solids in the treatment of this loading can then have an impact on digestion loading. This cycle will continue until the whole process has reached a "steady-state." These loading factors need to be accommodated in the consideration of large capital projects for treatment plant facilities.

The dynamic simulation capability of whole-plant process models presents a major benefit. With the advent of increasingly complex control strategies, such as ammonium-based aeration control (ABAC), dynamic whole plant process models can guide assessments of dynamic process response and help develop appropriate operational control strategies. As shown in Figure 1, a process model can be used to simulate dynamic airflow and ammonium response by inserting a

proportional-integral-derivative (PID) controller loop into the model. This can provide valuable information that can be used during design, startup, and optimization of an ABAC system.

Figure 1: Example of how a whole plant process model can be used to assess the impact of diurnal ammonium concentrations and required DO setpoints and control loop timing in ammonia-based aeration control (ABAC) loop optimization.

Whole-plant process modeling also provides needed insight into biological nutrient removal control optimization. A well-calibrated process model can complete thousands of simulations using different loading conditions and different potential operational configurations for a facility. For example, if a facility is looking to optimize solids retention time (SRT), return activated sludge (RAS) pumping, internal mixed liquor recycle (IMLR) pumping rates, and dissolved oxygen (DO) setpoints, there would be 24 (or more) potential combinations of configurations for operation. Using a process model and a probability distribution function of influent conditions, literally thousands of simulations can be completed on the 24 different configurations and the relative stability of the different configurations can be calculated. As shown in Figure 2, this can be used to identify the most stable configurations options without requiring years of operational field testing. Often, the modeling can be used to identify the "best few," and these can be further narrowed through field testing.
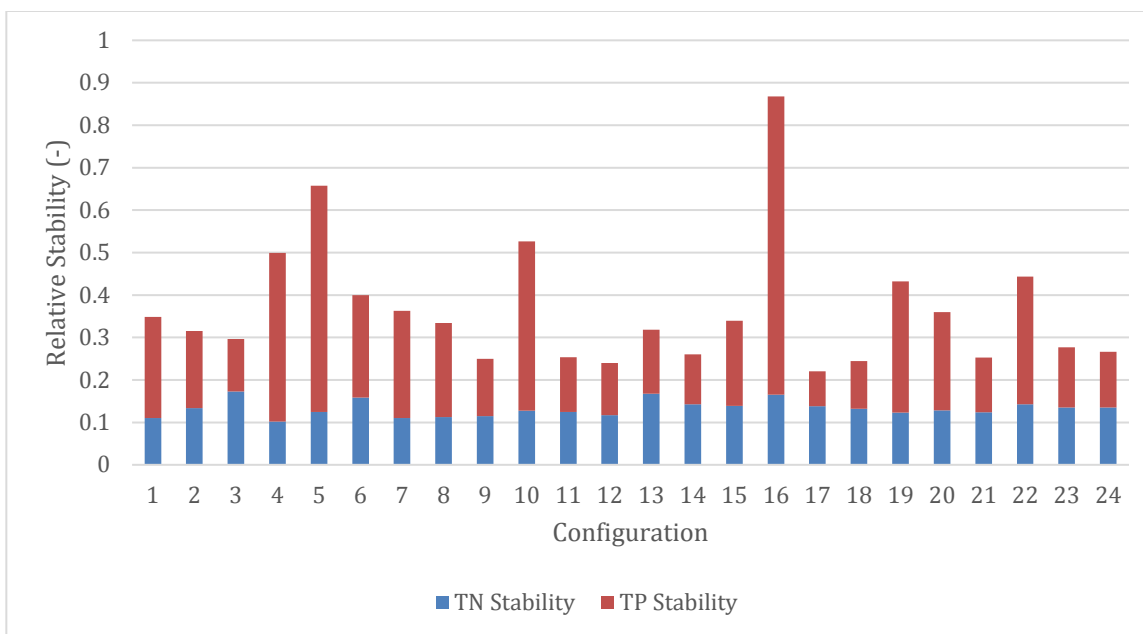
Figure 2: Process models provide the ability to run thousands of simulations related to reactor configuration for biological nutrient removal; as shown above, the stability of total nitrogen (TN) and total phosphorus (TP) can be assessed to identify the most stable operating strategy for existing infrastructure configurations.

The whole-plant commercial model might also be utilized for training and developing plant operations staff and might serve an organization as a succession planning tool. A model as a simulator would allow an operator to view the effect of changing operational parameters without actually impacting operations. Development of this potential was described in "Modeling Good Practices," March, 2017, WE&T, Snowling et al.

The use of whole-plant models for operator training and day-to-day operations assistance must be carefully planned and scoped. The main challenges facing utilities for application of whole-plant process models for daily operations are a lack of funding and staff training; data management; and confidence in model predictions. Often, the key process variables required for day-to-day operation can be developed using spreadsheet-based analyses and simplified calculations. An example of this alternative using quick, easy tools will be described further.

PROCESS CALCULATOR AS AN ALTERNATIVE
When it comes to making critical decisions, plant operators frequently find themselves in the "hot seat." One situation involves equipment failure that requires taking aeration tanks or final clarifiers out of service for maintenance. To an operator, it seems as though these situations occur when settling is slow and there are other operational challenges such as high flows. How does the operator make a decision as to whether tanks can be removed from service and avoid effluent violations?

The traditional method of determining the need for day-to-day process manipulation is operator intuition based on data, training, and experience. This technique is highly dependent upon the

operator to provide a course of action; different levels of training and experience can yield different results. An experienced operator decision may provide the best action if it is part of his/her knowledge base. However, for more complex decisions, where operating parameters are outside a typical range, for a sudden change in loadings, or where there are other unknowns, it may be more effective to rely on a more science-based method to bridge the gap between strict operator experience and the use of a commercial process model.

A process calculator would use a simpler set of equations than a commercial model to help make decisions. A process calculator with these characteristics was developed to help answer questions for the NSWTP operated by the Madison Metropolitan Sewerage District (MMSD or District) in Madison, Wisconsin. The calculator was programmed into the supervisory control and data acqusition (SCADA) system for this utility. The Madison process calculator uses readily available plant laboratory data in combination with continuously available SCADA process data to help answer operational questions. The system has been continuously calculating and providing data for real-time decision making since the 1990s. Using a SCADA system based process calculator can open up the additional possibility of using a data historian trending package to graph changes and results over time. An image of the operator interface is shown in Figure 3. A spreadsheet version was adapted for use at other much smaller facilities using the same basic equations and is shown in Figure 4.

Madison Process Calculator Details
A kinetic expression developed and generally utilized for a complete mix activated sludge system with cellular recycle is:

$$(1) \qquad X = \Theta_C/\Theta * Y (S_0 - S)/(1 + K_d \Theta)$$

Where

X = *mixed liquor concentration, (mg/L);*
Θ_c = *SRT, (days);*
Θ = *hydraulic retention time = $Vol_{(reactor)}/Q_{inf}$, (days);*
$Vol_{(reactor)}$ = *Reactor Volume, (m^3);*
Q_{inf} = *influent flow, (m^3/day);*
Y = *yield coefficient;*
S_0 = *initial substrate concentration, (mg/L);*
S = *final substrate concentration, (mg/L); and*
K_d = *decay coefficient, ($days^{-1}$).*

Two operational parameters normally tracked in treatment plant operation are F:M and SRT. Equation 1 can be reduced to the following using typical operational parameters and with the assumption that the effluent $BOD_5 = 0$ mg/L:

$$(2) \qquad 1/SRT = Y (F:M) - K_d$$

Where

SRT = *$Vol_{(reactor)}$ * MLTSS * 0.001 / (kg wasted/day), (days);*
$F:M$ = *(Q_{inf} * $BOD_{5(inf)}$) / ($Vol_{(reactor)}$ * MLVSS), ($days^{-1}$);*

MLTSS = mixed liquor total suspended solids, (mg/L);
BOD$_{5(inf)}$ = influent (primary effluent) 5-day biochemical oxygen demand, (mg/L);
and MLVSS = mixed liquor volatile suspended solids, (mg/L).

Figure 3: Inputs to and Outputs from the Madison, Wisconsin Process Calculator.

INPUTS

Target Sludge Age	15	Days
Temp	15	deg C
Inf BOD avg	250	mg/l
Inf TSS	250	mg/l
Inf NH3 avg to Aeration	40	mg/l
Effl TSS	35	mg/l
MLSS SVI	95	
MLSS % Volatile	75%	
Primary % Volatile	80%	
Inf Flow avg daily	0.76	Ml/D
Inf Flow last hour	0.95	Ml/D
RAS Flow avg daily	0.45	Ml/D
RAS Flow last hour	0.57	Ml/D
Aeration Tank Vol	0.61	Ml
Clarifier Area	74.6	Sq M
Primaries	1	Yes=1/No=0
% TSS Removal	60%	%
%BOD Removal	30%	%
Sludge Thickening		
Thickened Primary	5%	%
Thickened Waste	5%	%
Digestion	1	Yes=1/No=0
%VS Reduction	56%	%
Thickened Digested	5.5%	%

OUTPUTS

Inf BOD to Aeration	175	mg/l
Est Avg MLSS	3143	mg/l
Est Avg RAS conc	8382	mg/l
Hourly RAS conc	8382	mg/l
Waste MLSS Flow (zero TSS)	30398	Ml/D
Waste RAS Flow (zero TSS)	11399	Ml/D
Waste Sludge Mass	112	Kg/D
Waste MLSS Flow (adj for effl TSS)	23167	Ml/D
Waste RAS Flow (adj for effl TSS)	8688	Ml/D
Calc Hydraulic Loading for Hourly Flow	12687	L/sq m/D
Calc Solids Loading for Hourly Flows	74	Kg/sq m/D
Calc Hydralic Underflow Rate	7612	L/sq m/D
Max Clarifier Ldg for Hourly Flows	89	Kg/sq m/D
% of Clarifier Capacity Utilized	84%	
Aerobic Detention Time	19.3	hours
Total SRT	142.9	days
Required Nitrification Rate	0.00088	Kg NH3-N/Kg MLVSS/Hr
Max Rate for Temperature	0.0024	Kg NH3-N/Kg MLVSS/Hr
(0.003 lbs NH3-N/lb MLVSS/Hr @ 20 deg C)		

Sludge Storage

180 days storage	0.417	Ml
Mass of Raw Sludge	1090	Kg/D
% Volatile	78%	%
Raw Sludge Flow/day	3592	L/D
Mass of Digested Sludge	1062	L/D
Flow/day (with digester)	2316	L/D

Figure 4: Spreadsheet version of Process Calculator Used for a Smaller Community.

The Yield, (Y) and decay coefficient (K_d) based on BOD_5 removal and mixed liquor solids are not technically the same as the yield coefficient and decay coefficient defined in Equation 1, but fundamentally define the same relationship between biological growth and influent substrate.

The "pseudo" kinetic parameters of Equation 2 can be used to predict the resulting mixed liquor concentration for a given biological system tank volume, for a desired SRT and influent BOD_5 loading and can be used for estimating purposes regardless of the configuration as complete mix, plug flow, or other configurations.

Between 1984 and 1988, and prior to biological phosphorus removal, the NSWTP transitioned from partial nitrification and a 5-day sludge age to full nitrification at a 10 to 12-day sludge age. From Equation 1, the parameters "1/SRT" vs "F:M" were charted for the time period from 1984 to 1988 and an estimate for average overall "Y" and "K$_d$" were made from this analysis. For Madison, values of Y = 1.2 and Kd = 0.5 were determined with primary settling. These values have since seemed not only to provide reasonable estimates for Madison, but also for other plants treating domestic wastewater with primary settling before secondary aeration.

By expanding Equation 2, the following can be used to predict the average plant mixed liquor concentration.

(3) $\textbf{MLTSS}_\textbf{Est} = \textbf{(Y * Q}_\textbf{inf} \textbf{ * BOD}_\textbf{5}\textbf{) / (Vol}_\textbf{(reactor)} \textbf{ *}$
$\textbf{(1 / SRT + K}_\textbf{d}\textbf{) * Fraction Volatile)}$

Where
 MLTSS$_{Est}$ = Estimated MLTSS, (mg/L);
 Vol$_{(reactor)}$ = Reactor volume, (m^3); and
 Fraction Volatile = MLVSS/MLTSS.

Based on a mass balance for an activated sludge process without internal recycle flows, the estimated RAS concentration for any time period can be calculated as:

(4) $\textbf{RAS}_\textbf{Est} = \textbf{(Q}_\textbf{inf} \textbf{ + Q}_\textbf{ras}\textbf{) * MLTSS}_\textbf{Est}\textbf{/Q}_\textbf{ras}$

Where
 RAS$_{Est}$ = Estimated RAS concentration, (mg/L); and
 Q$_{ras}$ = Return Activated Sludge Flow, (m^3/d).

Estimates for daily average waste activated sludge (WAS) flow rates required for wasting of either mixed liquor (Qw$_{mlss}$) or return sludge (Qw$_{ras}$) to achieve a desired SRT can be made from the following general equations, respectively, assuming zero secondary effluent total suspended solids (TSS).

(5) $\textbf{Qw}_\textbf{mlss} = \textbf{Vol}_\textbf{(reactor)} \textbf{ / SRT}$

Where
 Qw$_{mlss}$ = Waste Activated Sludge flow of MLTSS, (m^3/d).

(6) $\textbf{Qw}_\textbf{ras} = \textbf{Vol}_\textbf{(reactor)} \textbf{ * Q}_\textbf{ras} \textbf{ / (SRT * (Q}_\textbf{ras} \textbf{ + Q}_\textbf{inf}\textbf{))}$

Where
 Qw$_{ras}$ = Waste flow of RAS, (m^3/d).

With the estimated mixed liquor concentration from Equation 5 above and the estimated waste sludge flow, an approximation of solids production can be made:

(7) $\textbf{WAS}_\textbf{Est} = \textbf{Qw}_\textbf{mlss} \textbf{ * MLTSS}_\textbf{Est} \textbf{ * 0.001}$

Where

WAS$_{Est}$ = Waste Activated Sludge Mass, (kg/d).

An adjustment can be made in solids production estimates for mixed liquor temperature. Growth in the winter is slower and WAS production higher. The modified yield coefficient, normally 1.2 for a standard temperature of 20°C, is adjusted by Equation 8 for other process temperatures. For the Madison, Wisconsin process calculation, the temperature utilized is the plant effluent temperature continuously recorded in the SCADA system.

(8)   Yt = Y$_{20}$ /1.01$^{(T-20)}$

Where

Yt = Yield coefficient at process temperature;
Y$_{20}$ = Yield coefficient at 20°C; and
T= Process temperature, (°C).

For assisting an operator in finding the proper daily wasting volume, the waste flow calculation can be modified to take into account effluent TSS as shown in Equation 9 (for wasting mixed liquor) and 10 (for wasting RAS) below.

(9)   Qw$_{mlss(adj)}$ = (Qw$_{mlss}$ * MLTSS$_{Est}$ − Q$_{inf}$ * TSS$_{effl}$)/MLSS$_{Est}$

Where

Qw$_{mlss(adj)}$ = Adjusted WAS flow of MLTSS, (m^3/d); and
TSS$_{effl}$ = Total Suspended Solids in Effluent Flow, (mg/L).

(10)   Qw$_{ras(adj)}$ = (Qw$_{ras}$ * RAS$_{Est}$ − Q$_{inf}$ * TSS$_{effl}$)/RAS$_{Est}$

Where

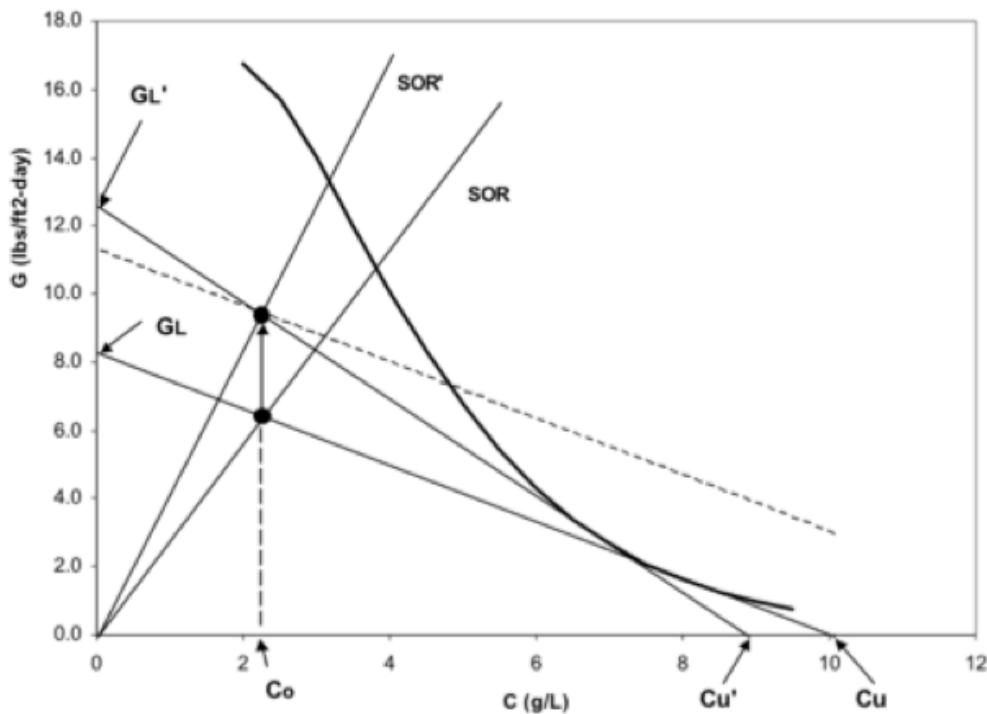Qw$_{ras(adj)}$ = Adjusted waste flow of RAS, (m^3/d).

Since the Madison, Wisconsin plant uses a variation of the University of Cape Town (UCT) process as shown in Figure 5, the solids concentration in the anaerobic zone is less than in the anoxic and aerobic portions of the tank and depends on the recycle ratio. The anaerobic zone volume is 1/6 of the total tank volume. The calculations for waste sludge flow and influent solids concentration to the clarifiers are modified in the mass balance for Madison to take the different zone volumes and concentrations into account in the equations above.
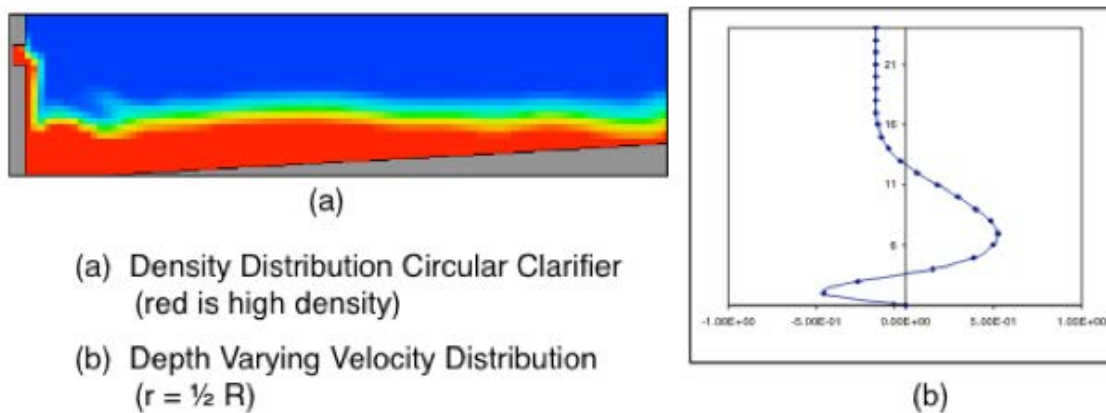
MADISON PROCESS SCHEMATIC
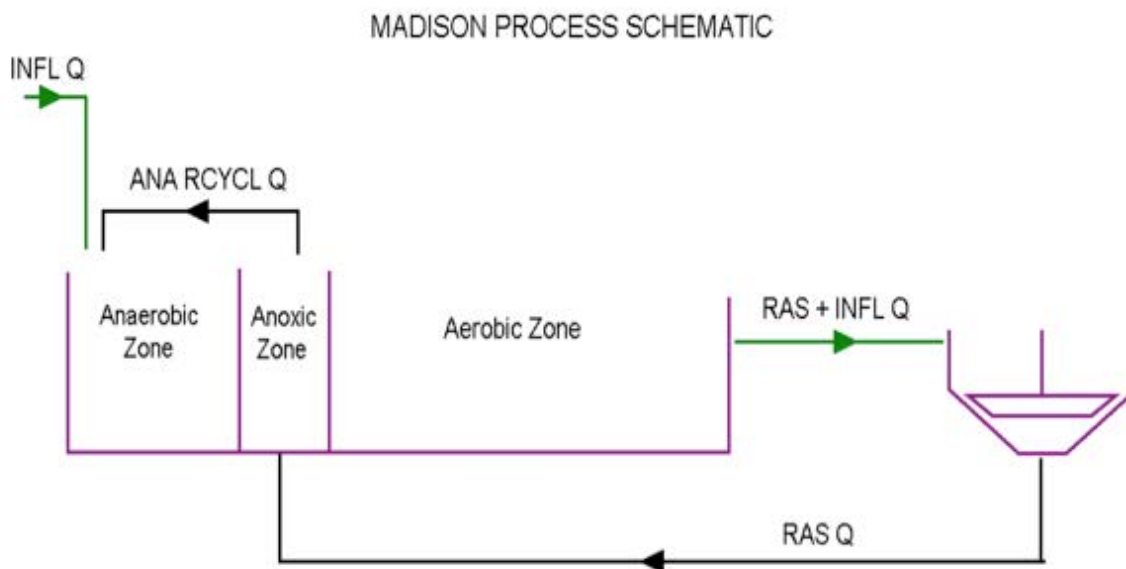
Figure 5: Madison Biological Nutrient Removal Configuration (Modified University of Cape Town Process).

Predicting Clarifier Capacity In the Process Calculator
Once the mixed liquor concentration is predicted and RAS rates are known, the clarifier capacity can be estimated based on sludge settling characteristics, solids loading, and clarifier surface area. Theory for this prediction has been researched for more than three decades and is termed solids flux theory. One of the most common analysis methods is "state point analysis," and is depicted graphically in Figure 6. Other methods have been developed utilizing computational fluid dynamics and numerical analysis. A typical output of computational fluid dynamics modeling is shown in Figure 7. Both state point analysis and computational fluid dynamics require determination of sludge settling velocities with stirred batch settling tests run at different dilutions. A test history is developed from a set of diluted batch settling tests and is only valid for the current bacterial populations and settling characteristics.

Figure 6: Settling Velocity Curve with State Point Analysis Lines.

(a) Density Distribution Circular Clarifier (red is high density)

(b) Depth Varying Velocity Distribution (r = ½ R)

Figure 7: Output from Computational Fluid Dynamics Modeling.

The alternative used for the Madison process calculator utilizes the standard 30-minute settleometer test results performed each day by plant operators combined with a "Design and Operating Diagram" researched and developed by Daigger and Roper (1985) for Milwaukee Metropolitan Sewerage District, Wisconsin. These relationships were later updated by Daigger (1995) using a larger database. The graphical relationship from the 1985 paper is shown in Figure 8. The operating diagram shown was generated by using settling velocity curves

associated with different sludge volume index (SVI) values, and then analyzing the solids loading capacity by using state point analysis for defining a maximum solids loading capacity associated with a given underflow rate and associated underflow solids concentration.

The data from the 1985 operating diagram was fit for Madison's process calculator with a series of logarithmic equations and then mathematical interpolation used to estimate between the curves for any given SVI. The approximate fits are illustrated in Figure 9. The fitted equations provided a method for quickly calculating the current percent utilization of the maximum clarifier solids loading capacity. The maximum solids loading capacity is iteratively calculated from the curves at the current RAS rate and SVI. This calculation is performed in the Madison SCADA and continuously predicts clarifier capacity as a percent of the maximum solids loading capacity during normal diurnal variations, as well as during storm flow conditions. The output data from this system can be recorded for future reference using a data historian or other data logging methods.
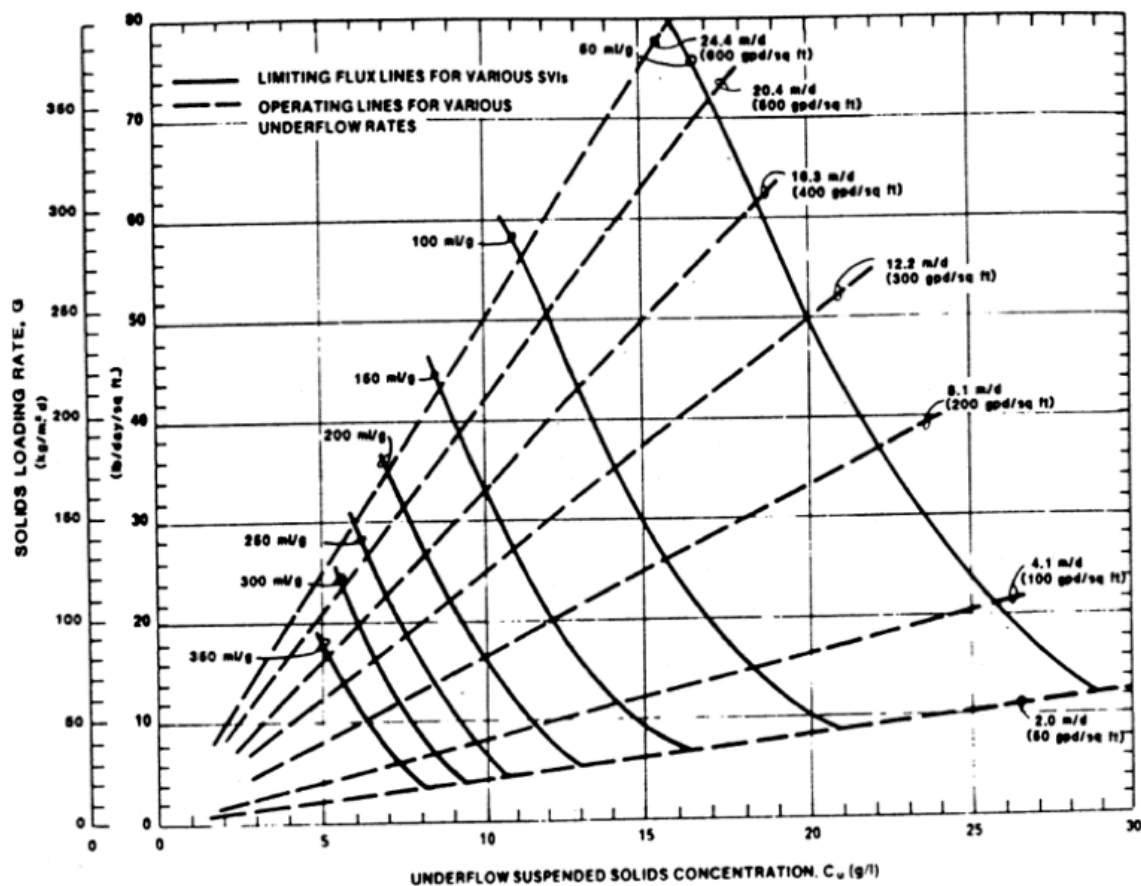
From Daigger and Roper, (1985), "The Relationship Between SVI and Activated Sludge Settling Characteristics", JWPCF, 57, p.859.

Figure 8: Clarifier Design and Operating Diagram.
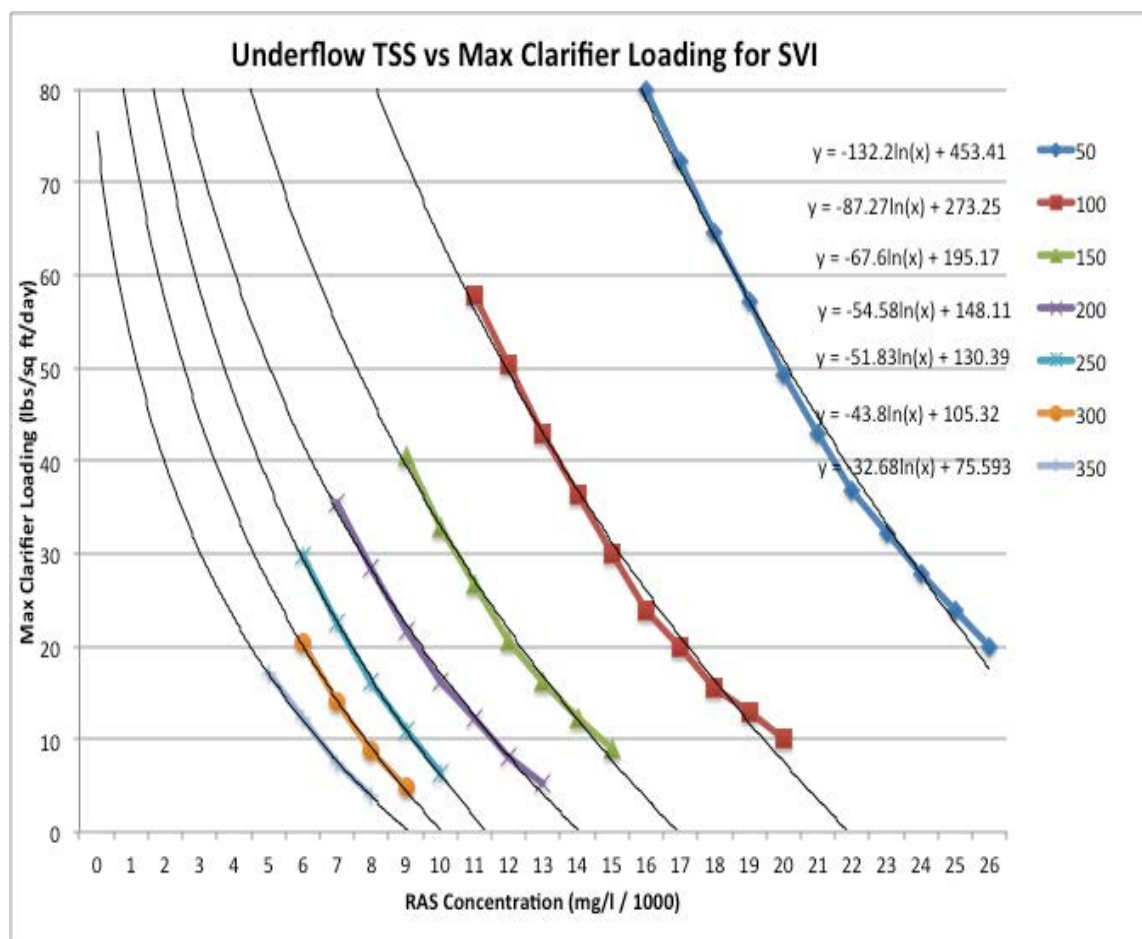
Figure 9: Curve Fit of Clarifier Design and Operating Curve.

Aeration System Calculations

The District has used fine pore ceramic diffusers for aeration since 1978. Most of the diffusers are ceramic discs originally installed between 1983 and 1985. After installation, the District worked extensively with the University of Wisconsin-Madison investigating the efficiency of the system using an off-gas testing apparatus. From this research, alpha standard oxygen transfer efficiency (αSOTE) in plug flow aeration tanks has been generally defined based on position in the tank. αSOTE is the standardized oxygen transfer efficiency (OTE) at zero DO and 20 °C in dirty process water conditions. Alpha (α) is a ratio from 0.0-1.0 and increases throughout the length of plug flow reactors as the waste is stabilized. α has been found to vary from about 0.3 to 0.7 in plug flow tanks and αSOTE from about 8% to 20%.

The actual oxygen transfer rate also depends on DO concentration. The transfer rate is inversely proportional to the DO deficit. For Madison's 16' deep aeration tanks the average maximum saturation concentration is 10.6 mg/L. So an estimated overall transfer rate can be made with the following equation:

(11) Field Oxygen Transfer Rate =
Process Air Flow Rate * Density of Air * 0.21 O₂ fraction *
αSOTE * (10.6 - DO$_{reactor}$) / DO$_{reactor}$

Where

Field Oxygen Transfer Rate, (kg/min)
Process Air Flow Rate, (m^3/min);
Density of Air, (kg/m^3);
αSOTE = Alpha Standard Oxygen Transfer Efficiency;
DO$_{reactor}$ = DO concentration in reactor, (mg/L).

The SCADA system uses real-time air flows rates and DO concentrations of reactors combined with historical αSOTE values to estimate the mass of oxygen currently being transferred. This mass can be compared to the estimated oxygen demand using empirical equations based on influent flow, BOD₅, total Kjeldahl nitrogen (TKN) and effluent nitrate values. The mass transferred can also be compared to an estimated maximum oxygen transfer rate capacity based on diffuser densities, αSOTE values, and blower air flow capacity. This can be used to estimate and record the percent of the maximum aeration capacity being utilized in operations.

Use of the Process Calculator
The process calculators shown in Figures 3 and 4 require the operator to enter lab data such as influent BOD₅, TKN, effluent TSS, and daily SVI data. The plant SCADA system continuously updates the number of tanks in service, average daily and hourly flows, as well as mixed liquor temperature (from plant effluent). From this data, the estimated mixed liquor suspended solids (MLSS) and RAS concentrations are estimated as described, estimated wasting flows are calculated, and the estimated percent of clarifier capacity computed.

Laboratory data for plant MLSS and RAS can be compared with the estimated values predicted by the calculator and differences investigated. Discrepancies at times have been indicators or alerts used to find what turned out to be valving or wasting problems. After correction, the solids concentrations of process reactors would return close to the predicted values of the calculator.

An important function of the process calculator is to allow the plant operator to check alternative process conditions. The calculations obtain data from the control programs, but operate independently of process control algorithms. The operator can freeze the data communicated from SCADA selecting "NO" to the "Program Entered" question on the display. The operator is then free to change any parameter manually to evaluate "what if" scenarios. So in addition to influent BOD₅ and SVI, the operator can change the number of clarifiers or aeration tanks in service, and input different influent and RAS flow rates to evaluate alternative conditions that might occur during plant maintenance, higher flows, a higher SVI, or the capacity that might be associated with future loading conditions.

Checking clarifier capacity prior to taking tanks out of service has been one way the manual functionality of the calculator has been utilized. This can be demonstrated by comparing Figures 10 & 11. Figure 10 is an exploded view of the current actual operating data shown in Figure 3, while Figure 11 shows the operator turning off automatic feedback of data and simulating the removal of one of two aeration tank from service and reducing the influent flow by three percent.

The results show the clarifier capacity exceeded, with the clarifier capacity going from 52% to 112% utilization. Considering the result, the operator would evaluate options such as shifting flow to other available treatment units or significantly increasing the return sludge rate.

The calculations were extremely useful in Madison prior to 1997 and the implementation of biological phosphorus removal; the SVI frequently varied and would often exceed 200. The effect of high flows, taking aeration tanks or clarifiers out of service, and predicting of the effect of changing SRT and wasting rates often required quick and accurate prediction considering the current sludge settling quality.

Figure 10: Exploded view of Plant 3 current operational data.

Figure 11: MMSD Process calculator with "NO" for program entered values from Plant 3. The changes simulate removing one aeration tank from service and decreasing the flow 3%. Outputs show the clarifier capacity would be exceeded, 112% of maximum capacity.

DISCUSSION: WHEN TO APPLY A WHOLE PLANT PROCESS MODEL VS. A PROCESS CALCULATOR

As with most things in our industry, there are no clear answers when it comes to the tool to use for operations and design. The process calculator is not a substitute for process modeling in capacity assessment and design, particularly for nutrient removal predictions. Process models also provide particular power for assessing different operating configurations and control strategies.

Madison Metropolitan Sewerage District has actively used both whole plant process models and the process calculator for operations and decision making over the past 24 years. The District first relied on process models for decision making in the mid-1990s. In 1994, the District began operating two pilot plants to evaluate biological phosphorus removal options to meet pending phosphorus limits. The UCT and Anaerobic-Anoxic/Oxic (A^2O) processes were piloted for 1½ years to evaluate feasibility. The results of the piloting were used to calibrate a BioSim™ process model (a predecessor product to BioWin™), which was then used to evaluate many other process alternatives. The modeling showed another alternative, UCT without mixed liquor recycle (modified UCT), would be the most efficient and cost effective for meeting the pending limits. The pilot plant was re-started and operated for an additional six months to verify the recommended process option and the design parameters were confirmed and later implemented.

Similarly, in 2012, MMSD was planning for a much more restrictive phosphorus limit and potential nitrogen limits. In a report completed prior to facility planning, Pro2D™ modeling was used to analyze best options for nine different effluent nutrient scenarios shown in Table 1. The modeling was critical for efficiently defining treatment options for different effluent limits. Modeling was essential for narrowing down many potential options, estimating costs, and determining the course of follow on facilities planning.

TABLE 1: Nutrient Discharge Limit Scenarios

Scenario	Total Phosphorus, mg/L	Total Nitrogen, mg/L
1	0.225	None
2	0.130	None
3	0.075	None
4	0.225	10
5	0.130	10
6	0.075	10
7	0.225	3
8	0.130	3
9	0.075	3

Currently, the District is relying on BioWin™ process modeling as part of its toolset to evaluate innovative aeration control strategies to achieve more energy-efficient nitrogen and phosphorus removal. Process simulation work is being combined with pilot scale work to develop innovative operational strategies related to nitrite shunt, ammonium-based aeration control, and improved biological phosphorus removal.

The District has also purchased a modeling program to independently analyze options, with the idea that the model might also be used for future simulation training for operators as suggested in "Modeling Good Practices," March, 2017 WE&T article. Having this capability available on-site has not diminished the functionality available for operations-related decisions in the SCADA process calculator. The purpose of a process calculator is to provide a familiar and fast tool for operator decision-making and training. The process calculator can also be readily utilized for much smaller facilities in a spreadsheet format as shown in Figure 4.

Table 2 illustrates that there is a time and a place for each type of modeling system. The key to effective utilization of these is to understand the question being asked and to use the simplest tool to answer that question. There is naturally some overlap in the functionality available, and it could be anticipated that as whole-plant process models are further developed, they may fill some of the needs that a simpler process calculator can now satisfy. This may require that a "simplification" mode for software be developed to enable an operator to more easily assess process problems in a way that is available now with a process calculator.

TABLE 2: A time and a place for all models

Function	Whole-Plant Process Model	Process Calculator
Daily WAS pumping requirements		+
Daily check on mixed liquor concentration		+
Clarifier capacity check for varying operating scenarios	+	+
Secondary clarifier sludge blanket rise	+	+
Estimate of wasted mass and storage voume needs	+	+
Operator Training	+	+
Biological nitrogen and phosphorus configurations	+	
Aeration control loop troubleshooting	+	
Projecting future aeration requirements	+	
Daily aeration basin blower adjustments		+
Tracking aeration capacity utilization		+
Solids processing polymer dosing adjustments		+
Daily chemical dosing adjustments		+
Chemical dose optimization	+	

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