Estimating oxygen transfer capacity of a full-scale pure oxygen activated sludge plant

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ABSTRACT: A process-water, oxygen transfer compliance test was performed in November, 1983 on a 6.0-m^3/s (138-mgd) high purity oxygen activated sludge plant. The plant failed this and a subsequent process water test and the failure required the development of a procedure to determine oxygen transfer capacity of the plant. The American Society of Civil Engineer's clean water oxygen transfer standard was used in conjunction with process modeling and pilot-scale alpha factor testing. Clean water test results and a dynamic process model which predicts head-space gas purity are presented. *J. Water Pollut. Control Fed.*, **61,** 208 (1989).

KEYWORDS: oxygen transfer, aeration, oxygen, activated sludge, wastewater, modeling, simulation.

In 1973, the Sacramento Regional County Sanitation District was formed to provide wastewater collection and treatment for 17 separate agencies and industries that operated 21 separate treatment plants. To reduce costs and improve efficiency, a new regional plant was required. Planning and design for this new plant began in 1973, with construction beginning in 1976. The regional plant now treats all major sources of wastewater generated within Sacramento County.

The design engineers selected the high-purity oxygen activated sludge process for the regional plant. The selection was based on the successful process performance of a pilot plant study conducted at the city of Sacramento's main treatment plant during the last 6 months of 1973. Other reasons for selecting the process were the concerns for combined municipal and industrial wastewater treatment and odor control. Because the regional plant would have to treat a significant amount of seasonal food processing (canning) wastewater, it was believed that a high-purity oxygen system would perform better.

Plant startup began in November, 1982, and the first oxygen dissolution system compliance test was performed in November, 1983. This test was performed to verify that the specified amount of oxygen could be transferred at or below the specified power consumption rate. The compliance test was performed on process water during plant operation, as opposed to "clean water" in the conventional way. When the treatment process was designed and the specifications written, the American Society of Civil Engineers (ASCE) standard[1] for clean water testing did not exist.

The first process water test failed to demonstrate the specified performance. Two years later a second process water test showed the same result. For 2.5 years, the process, with respect to its mass transfer capability was investigated. The county and its consultants developed clean water data and a model to verify the process's oxygen transfer capability and shortcomings. The results of the model were eventually accepted by all parties. The plant model showed that the original oxygen transfer specifications, with the exception of an additional capacity requirement in the sludge reaeration mode (a form of step feed), could be met.

Plant Description

The Sacramento Regional Treatment Plant is a full secondary treatment facility providing treatment for 6.0 m^3/s (138 mgd), and includes raw and effluent pumping, primary clarification, secondary treatment with the high-purity oxygen activated sludge process, cryogenic oxygen production, disinfection, sludge thickening, and anaerobic digestion of waste sludges. The nominal design basis is shown in Table 1. The nominal design flow rate is 5.0 m^3/s (115 mgd) for dry weather, non-canning season, 6.0 m^3/s (138 mgd) for dry weather, canning season, and 10.5 m^3/s (240 mgd) for peak, wet weather conditions. The original oxygen transfer performance specifications were written using the 6.0-m^3/s flow rate, which will be used as the design flow rate throughout this manuscript, unless otherwise noted. Table 2 shows the secondary influent wastewater characteristics.

The plant has several unusual requirements. The effluent must be diverted to storage basins during periods when the Sacramento River velocity is less than 0.15 m/s (0.5 ft/sec). The stored effluent cannot be discharged directly to the Sacramento River, and must be returned to the plant influent.

An interesting aspect of this plant is its deep tanks and turbine aerators. The majority of high-purity oxygen plants use low-speed mechanical surface aerators. This plant uses turbine aerators that have a conical gas diffuser located 7.7 m (25 ft) below the liquid surface. High purity oxygen is normally released only in the first of four stages, but

Table 1—Nominal dry weather design basis for the Sacramento Regional Treatment Plant.

Process	Design basis
Primary clarification	Twelve primary clarifiers sized at 0.41 $m^3/m^2 \cdot d$ (1170 gal/sq ft/day)
Aeration basins	Eight trains of four basins (stages) in series, each measuring 14.6 m wide × 14.6 m long × 9.1 m deep (48 ft × 48 ft × 30 ft), providing a hydraulic retention time of 2.9 hours and an F:M ratio of 0.47
Oxygen production capacity	Two cryogenic oxygen plants, each producing 91 tonnes · d (100 tons/day) of 97% pure oxygen
Oxygen transfer capacity	Each train is equipped with four turbine aerators at 56 kW (75 hp), 45 kW (60 hp), 30 kW (40 hp), and 30 kW, with eight recirculation blowers totaling 1025 kW (1375 hp)
Final clarification	Sixteen 40-m (130-ft) circular tanks providing 0.23 $m^3/m^2 \cdot d$ (650 gal/sq ft/day) overflow rate.

can also be released in the other stages. Recirculation blowers are located in a central blower building. They take suction on the gas space of each stage, recirculate gas to the turbine diffusers, and are manifolded so that different blowers can be used for different stages. This unique feature gives added flexibility so that a wide range of gas recirculation rates are achieved. Normally, the gas in each stage is not mixed with other stages. To achieve the maximum mass transfer rate, high-purity oxygen is fed directly to Stage 1 recirculation blowers. It was the manufacturer's intent that the high-purity oxygen feed provide the entire gas flow to the Stage 1 blowers and turbines. The plant is designed so that the conventional and sludge reaeration modes can be used. Figure 1 is a schematic of the aeration basins.

Performance Warranty

As indicated earlier, a process water performance warranty was provided instead of a clean water specification. This was done partially because the ASCE clean water specification did not exist at the time the plant was designed, and in part because the oxygen transfer capability of a high-purity oxygen plant is strongly influenced by oxygen gas purity in each stage, which is not addressed by the ASCE standard or by clean water testing methods that existed when the plant was designed. Furthermore, the designer wanted to warrant other parts of the process, particularly the cryo plants.

The following process warranty was provided for the conventional or normal process mode and sludge reaeration modes provided that it was operating at the specified operations conditions:

- In the conventional mode, transfer 125 tonnes/d (138 tons/day) of oxygen with not less than 63.6 tonnes/d (70 tons/day) occurring in Stage 1, given 139 tonnes/d (153 tons/day) high-purity (97%) oxygen feed rate (90% oxygen utilization rate). It was further stipulated that this transfer occur at 6-mg/L average mixed liquor suspended solids (MLSS) dissolved oxygen (DO) concentration at a mixed liquor temperature of 28°C.
- Maintain an average DO in all stages of 6.0 mg/L or more, with no stage having less than 4.0 mg/L.
- Consume no more than 1600 kW (2144 hp). This total includes power for the turbine mixers and recirculation blowers, but excludes power associated with the cryo plants.
- In the sludge reaeration mode, transfer 160 tonnes/d (176 tons/day) at 177 tonnes/d (195 tons/day) of 97% purity oxygen feed rate. The temperature and DO concentrations for these requirements equaled those for the normal mode, and no maximum power was specified.
- In the event of non-compliance, the manufacturer was required to modify the system to meet the specified transfer rates. If the power consumption of the original or modified system exceeded 1600 kW, a power penalty of $3600/kW was to be assessed.

To test these warranty conditions, a full-scale process water test was planned. Oxygen transfer was estimated over a 7-day period using a steady-state material balance across the aeration basins. The material balance procedure required that the inlet and exit gas and liquid flow rates; inlet 5-day biochemical oxygen demand (BOD_5), chemical oxygen demand (COD), soluble COD, total suspended solids (TSS), volatile suspended solids (VSS), alkalinity, stage DO, and alkalinity concentrations; stage oxygen and CO_2 gas purities; and return sludge flow rate and concentration be measured at periodic time intervals, ranging from daily composites to instantaneous measurements every 4 hours. Stage gas purities were measured by collecting a sample from the gas head space of each stage; DO was measured by inserting a DO probe on a long shaft into the mixed liquor of each stage. Power was measured every 4 hours.

The COD, BOD, TSS, pH, and alkalinity data were not used in the material balance calculations for oxygen uptake rate (OUR). They were collected to ensure that the process met specified treatment efficiency, and that the influent wastewater met the design specifications shown in Table 2. This was necessary because influent wastewater characteristics can affect the gas space purity profile and oxygen transfer rates.

Table 2—Primary effluent water quality (design basis).

Parameter	Operating value
BOD_5, total	175 mg/L
BOD_5, soluble	114 mg/L
COD	335 mg/L
TSS	77 mg/L
VSS	62 mg/L
Temperature	28°C
Alkalinity (as $CaCO_3$)	160 mg/L
pH	7.1
Alpha factor	0.8
Beta factor	0.95

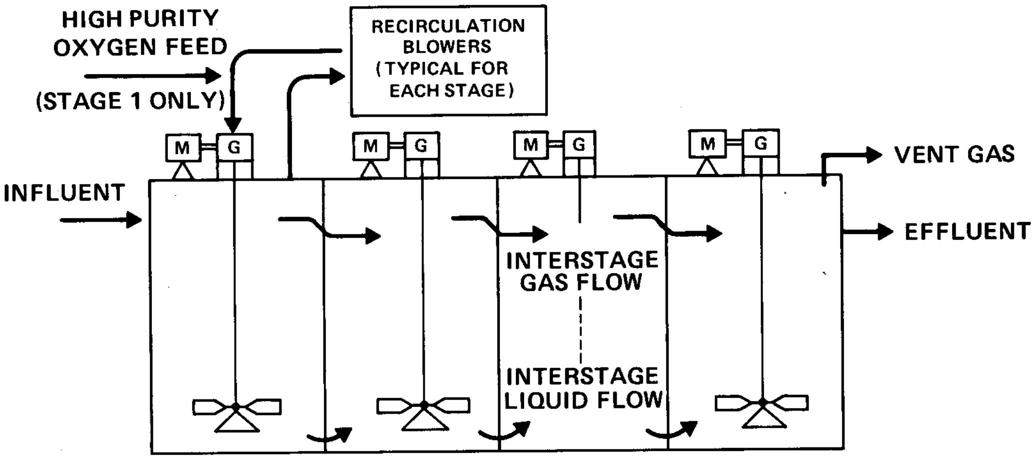

Figure 1—Plant schematic.

The data were reconciled by determining the oxygen transfer or OUR through direct measurement (input oxygen mass flowrate – the vent oxygen mass flowrate) and correcting this to the warranty conditions. The volumetric oxygen transfer coefficient, K_La, was estimated as follows:

$$K_La = \frac{OUR}{\alpha\theta^{T-20}[H\beta Y_oE_p - \text{DO}]} \tag{1}$$

Where

K_La = volumetric mass transfer coefficient, tonnes/d;
OUR = oxygen uptake rate, tonnes/d;
α = alpha factor, dimensionless;
θ = theta factor, 1.024, dimensionless;
H = Henry's law constant, g/m$^3\cdot$atm;
β = beta factor, dimensionless;
Y_o = oxygen gas pressure, atm;
E_p = effective pressure ratio, dimensionless;
DO = dissolved oxygen concentration, g/m^3; and
T = actual MLSS temperature, °C.

Alpha and beta factors were specified as 0.8 and 0.95, respectively, by the design engineer. The oxygen purity and DO concentration were measured in each stage of each train and averaged using a power-weighted ratio of each stage's mixer and blower power. These averages were calculated by multiplying each stage parameter by the total blower and turbine wire power. Products over all stages were summed and divided by the total power.

The effective pressure ratio was used to account for the hydrostatic pressure and was defined in the manufacturer's submittal as sparger mid-depth, which at 7.7 m depth is 1.37. Because the test was run for 7 consecutive days and DO and gas purity were measured at 4-hour intervals in each stage, it was necessary to evaluate Equation 1 forty-two times. The overall performance was evaluated as follows:

$$OUR = \alpha\overline{K_La}\theta^{28-20}[HY_o\beta E_p - \text{DO}] \tag{2}$$

Where

$\overline{K_La}$ = 7-day average volumetric mass transfer coefficient.

The 7-day average OUR defined by Equation 2 was the warranty oxygen transfer rate. The warranty or specified values of α, β, E_p, DO, and Y_o were used in Equation 2. Unfortunately, no procedures were specified to determine α and β. Consequently, an error in the α or β values in Equation 1 will bias the estimate of OUR in Equation 2. This cannot be corrected. Also, there was no specified method to explain differences in the specified Y_o and the measured Y_o.

It was thought that insufficient wastewater may exist to test all eight trains at full capacity at plant startup. Therefore, a provision was made in the specification that allowed fewer than eight trains to be used for testing. For example, if there was insufficient wastewater to produce 125 tonnes/d of oxygen demand, fewer trains could be used with a linearly proportional decrease in the required oxygen transfer and allowable energy consumption.

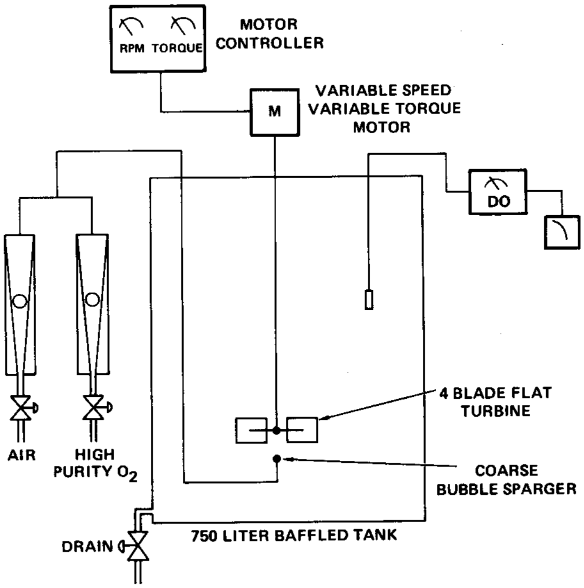

Figure 2—Alpha testing apparatus.

Table 3—Alpha factor test results.

Test series	Liquid	Date	Number of tests	K_La, min^{-1}	Alpha factor
1 S	Clean water	6/84	5	0.139 ± 0.005	
1 S	Stage 1	6/84	4	0.066 ± 0.019	0.48 ± 0.14
1 S	Stage 2	6/84	2	0.066 ± 0.008	0.48 ± 0.06
1 S	Stage 3	6/84	1	0.082	0.59
1 S	Stage 4	6/84	2	0.084 ± 0.002	0.61 ± 0.17
1 S	Primary effluent	6/84	2	0.054 ± 0.001	0.39 ± 0.008
1	Clean water	7/84	4	0.26 ± 0.007	
1	Stage 1	7/84	4	0.14 ± 0.007	0.55 ± 0.03
1	Stage 2	7/84	2	0.17 ± 0.001	0.67 ± 0.003
1	Stage 3	7/84	2	0.16 ± 0.001	0.62 ± 0.005
1	Stage 4	7/84	2	0.21 ± 0.012	0.79 ± 0.05
2	Clean water	7/84	3	0.17 ± 0.004	
2	Stage 1	7/84	2	0.0935 ± 0.003	0.54 ± 0.02
2	Stage 2	7/84	2	0.105 ± 0.001	0.60 ± 0.006
2	Stage 3	7/84	2	0.095 ± 0.001	0.55 ± 0.006
2	Stage 4	7/84	2	0.098 ± 0.002	0.56 ± 0.01

First Process Water Test

The first process water test was conducted November 2–9, 1983. Five trains were operated and an average of 42.5 tonnes/d were transferred using 889 kW. Directly scaling this transfer rate to eight trains gives 68 tonnes at 1422 kW. Oxygen utilization averaged 94.5%. The mass transfer rate was lower than expected and the manufacturer began to look for problems toward the end of the test period. This transfer was far short of the warranty conditions of 125 tonnes/d. Also, the DO in various stages did not meet the minimum measurement of 4.0 mg/L.

The manufacturer suspected that α was much less than 0.8. Consequently, a series of crude batch tests was performed in a 4-L vessel containing a fine pore stone diffuser. Primary effluent was used as the liquid for testing, as the design engineer's specification referenced an α value associated with primary effluent, as opposed to the mixed liquor. Alpha factors were then calculated by estimating K_La values from nonsteady-state reaeration of primary effluent spiked with mercuric chloride that terminated oxygen uptake. These K_La values were then divided by K_La values determined from reaeration of tap water that had been deoxygenated with nitrogen gas. A series of nine tests was performed. The average value for α was 0.35 with a minimum of 0.28 and a maximum of 0.42. Measurements for the determination of β were also taken; the average value was 0.95.

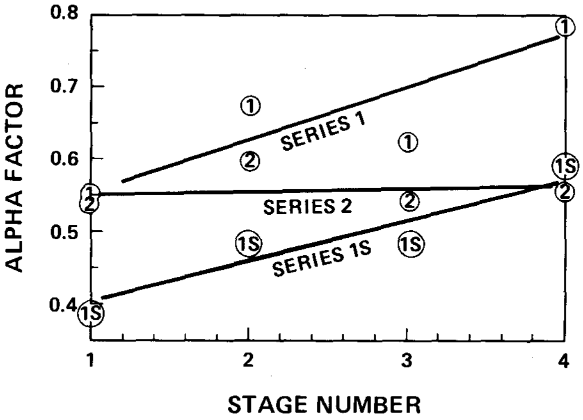

Figure 3—Alpha factor versus stage number.

The oxygen purity profile in the four stages during the process test was lower than anticipated. The manufacturer's analysis of the first process water test indicated an expected 66% power-weighted oxygen purity. The measured purity was only 52%.

When the values for α and β were incorporated into Equation 2, the warranty oxygen transfer increased to 115 tonnes/d for five trains in service, or 184 tonnes/d for eight trains, at 889 and 1422 kW, respectively. The manufacturer claimed that the wastewater and operating conditions during the test differed from those specified, and thus modified Equation 2 with an expected gas purity and α, changing the test conclusion from a 45% shortfall to a 47% excess in oxygen transfer capacity. The county and its consultants were unwilling to accept these calculation procedure modifications without documentation and verification.

Alpha factor testing. The first attempt to resolve the discrepancy in test result interpretation was to determine α. A test program was established in which primary effluent, mixed liquor, and a more appropriate apparatus were used. A realistic α for full-scale operation can only be determined using similar aeration devices. A fine pore stone in a 4-L bucket was convenient, but inappropriate, to determine α for the first process water test.

Figure 2 shows the apparatus used in this work. The 750-L aeration vessel was equipped with four baffles at 90-degree spacing. Each baffle was 10% of the tank diameter. The variable speed motor and gear box was selected to monitor rpm and torque, both of which change for different gas flow rates and water quality. A mixture of high-purity oxygen and compressed air was used. It was necessary to elevate the equilibrium DO concentration

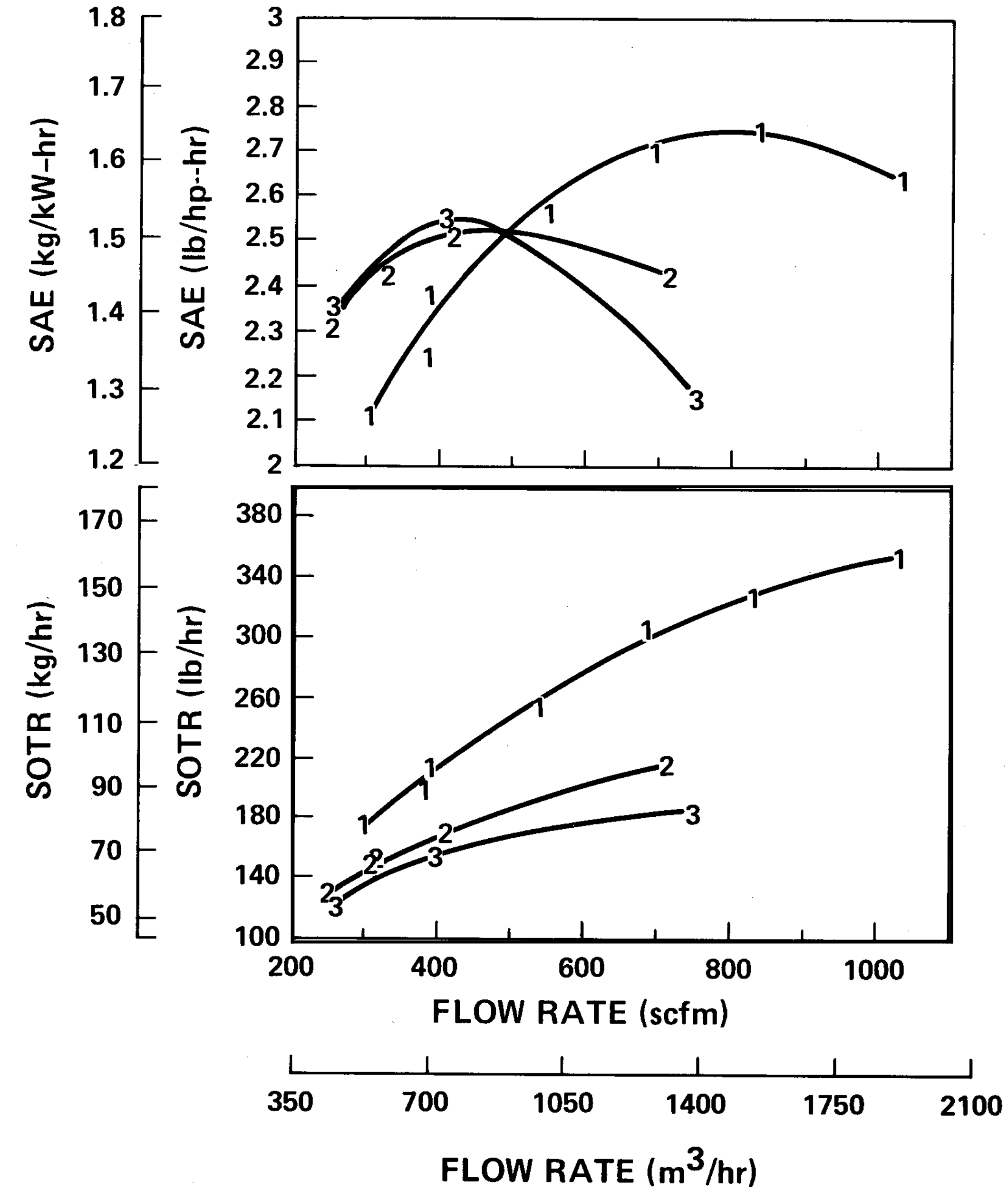

Figure 4—Clean water test results.

(C^*_∞), in the vessel in order to satisfy the oxygen uptake rate; otherwise, the DO in the test vessel would not change sufficiently to estimate K_La.

Researchers have recommended mercuric chloride or other chemicals that poison the mixed liquor to reduce its oxygen uptake rate to zero. This procedure was not used during this study, as there were concerns that the poisoning process might change the alpha factor. An alternate procedure that requires the oxygen uptake rate to be measured periodically during reaeration of a mixed-liquor sample was used.[3] Oxygen uptake rates were determined by collecting a sample from the 750-L tank, shaking if necessary to elevate DO concentration, taking a series of DO measurements, and then briefly recording the decline in DO concentration of the sample in a stirred BOD bottle.

In this way, a mathematical analysis procedure very similar to the ASCE nonsteady-state procedure can be used. To calculate K_La, the sum of squares was minimized as follows:

$$SS = \sum (DO_t - DO_t^o)^2 \qquad (3)$$

Where

DO_t^o = measured DO concentration at time t, and
DO_t = calculated DO concentration at time t.

The DO_t was calculated by integrating the following equation:

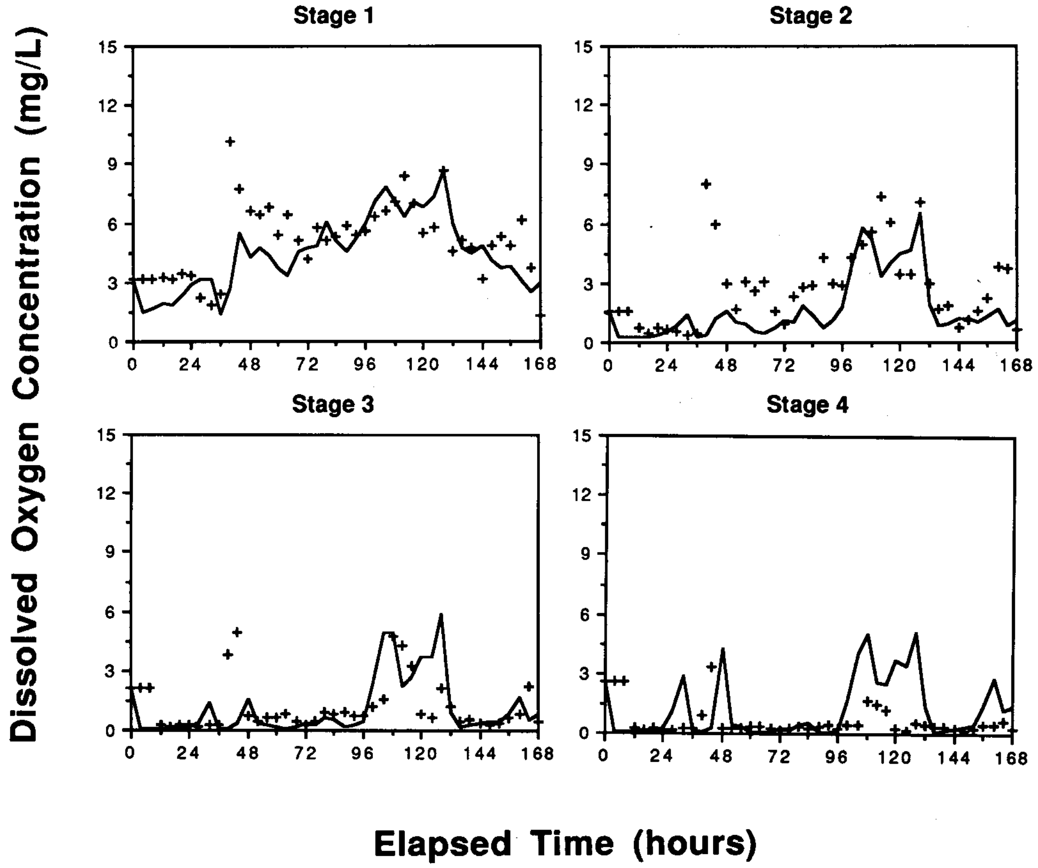

Figure 5—First process water test simulation: DO.

$$\frac{\mathrm{dDO}_t}{\mathrm{d}t} = K_La(C^*_\infty - \mathrm{DO}_t) - r(t) \qquad (4)$$

Where

C^*_∞ = equilibrium DO concentration, and
$r(t)$ = DO uptake rate as a function of t.

A procedure using the ASCE-supplied nonlinear least squares program to find the minimum in Equation 3 has been developed.[4] This technique requires that $r(t)$ in Equation 4 be adequately described by an exponential function. Another procedure, however, allows $r(t)$ to be an arbitrary function of time.[3] For this case, a second-order Lagrangian interpolation of the measured data points was used to model $r(t)$.

The apparatus in Figure 2 was placed on wheels and moved from stage to stage in order to determine α for each stage. A submersible pump, placed in each stage's sample port, was used to fill the tank. Filling time was kept minimal to keep the mixed liquor as fresh as possible, as mixed liquor in endogenous respiration generally has an elevated α.[2]

Table 3 presents results from the series of performed tests (PE = primary effluent and CW = clean water). Series 1S was performed with a 7.6-cm (3-in.) marine-type impeller at high rpm. In general, it was impossible to approximate the full-scale power density in the test tank using this impeller. A larger, 24-cm (9.5-in.) four-blade flat turbine was used later at lower rpm. This impeller, because it consumed more power, provided for conditions that were closer to those in full-scale tanks. These results are reported as Series 1.

Testing for Series 1S and 1 was performed between 9 a.m. and 3 p.m. This corresponds to the period of increasing plant load. Testing for Series 2 was performed in the early morning hours, between 5 a.m. and 10 a.m., which corresponded to the period of lower loading.

The trend for α is shown in Figure 3. Generally, α increases in the later stages. Some of the variability may be explained by loading changes, as it was impossible to perform all tests under the same plant load. During the periods when tests were performed, the approximate plant F:M ratio was 1.03 for Series 1S and 0.90 for Series 1 and 2. The corresponding mean cell retention times, calculated using sludge inventory in the aeration basins and secondary clarifiers, were 2.9 and 2.8 days, respectively. The power-weighted average α values (using the expected power consumption for the full-scale system) for Series 1S, 1, and 2 were 0.52, 0.64, and 0.56, respectively. The average α associated with primary effluent and the small impeller was 0.39 ± 0.01. This compares with primary effluent alpha values of 0.35 ± 0.07 that were determined by the manufacturer in the process water test.

The precision of the tests among replicates was quite good, but the absolute magnitude of α was much less than

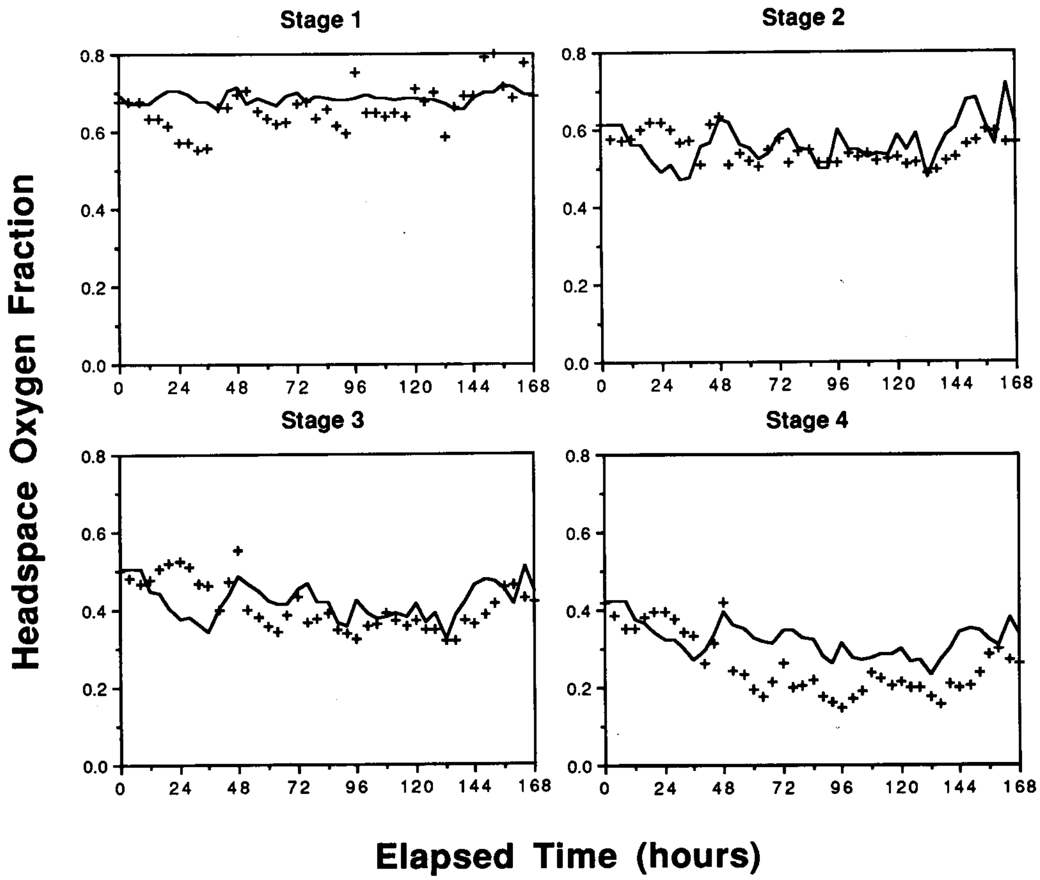

Figure 6—First process water test simulation: oxygen purity.

specified by the designer. The result did not resolve the dispute with the manufacturer, as a rather substantial power penalty and performance shortfall existed when the new α was factored in Equations 1 and 2, along with the measured gas purity of 52%. Furthermore, there was no way to show that the α-test apparatus accurately simulated the full-scale aeration system.

Clean water test. A hypothesis on the cause of the mass transfer rate deficiency was stipulated to be the specified clean water transfer efficiency of the turbines. The warranty specifications did not state the clean water transfer efficiency, but manufacturer's documents indicated a clean water transfer efficiency of 2.43 kg O_2/kW • h (4.0 lb O_2/hp/hr) in conventional mode, and 2.75 kg O_2/kW • h in sludge reaeration mode. The manufacturer used power units of brake and shaft horsepower, as opposed to wire horsepower. Clean water efficiency increased in the reaeration mode because gas recirculation rates increased and not because process conditions were different. Furthermore, an examination of documents supplied by the manufacturer revealed that the clean water transfer efficiency had been scaled up from 1.95 kg O_2/kW • h observed in their testing program in tanks 6.4 m deep to 2.43 kg O_2/kW • h in the county's 9.1-m deep tanks. Using the manufacturer's estimated blower and motor-gearbox efficiency, the standard oxygen transfer efficiency (or standard aeration efficiency, SAE), which is based upon wire horsepower, ranges from 1.83 to 2.05 kg O_2/kW • h (3 to 3.4 lb O_2/hp/hr) for the conventional and reaeration modes.

A clean water test was planned for Stages 1, 2, and 3. Stage 4 was not tested because it was identical to Stage 3. Train 8 was prepared for testing by draining and cleaning several times. This train had been previously used and was contaminated with mixed liquor. The openings between stages were blocked with plywood barriers. These barriers were designed to be opened and closed from the tank top using ropes and pulleys. During tank filling the barriers were opened to avoid damage from differences in hydrostatic pressure.

Train 8 was isolated from the high-purity oxygen system by closing the appropriate valves. Atmospheric air was supplied from a 150-hp positive displacement blower that was connected to each turbine through flexible hoses. To measure gas flow rate, a 12-m (40-ft) flow tube containing an orifice plate and a multiple-ported pitot tube was installed between the blower discharge and the turbines. Initially, there were severe problems in measuring flow rate accurately because the pitot tube's position along the flow tube influenced its flow indication. Also, the agreement with the orifice plate was poor. It seemed that there was some type of standing pressure wave in the air piping. After installing a noise silencer between the blower discharge and flow tube to function as a pulsation dampener, the problem was eliminated and the pitot tube measure agreed with the orifice plate measure to within ±1.5%.

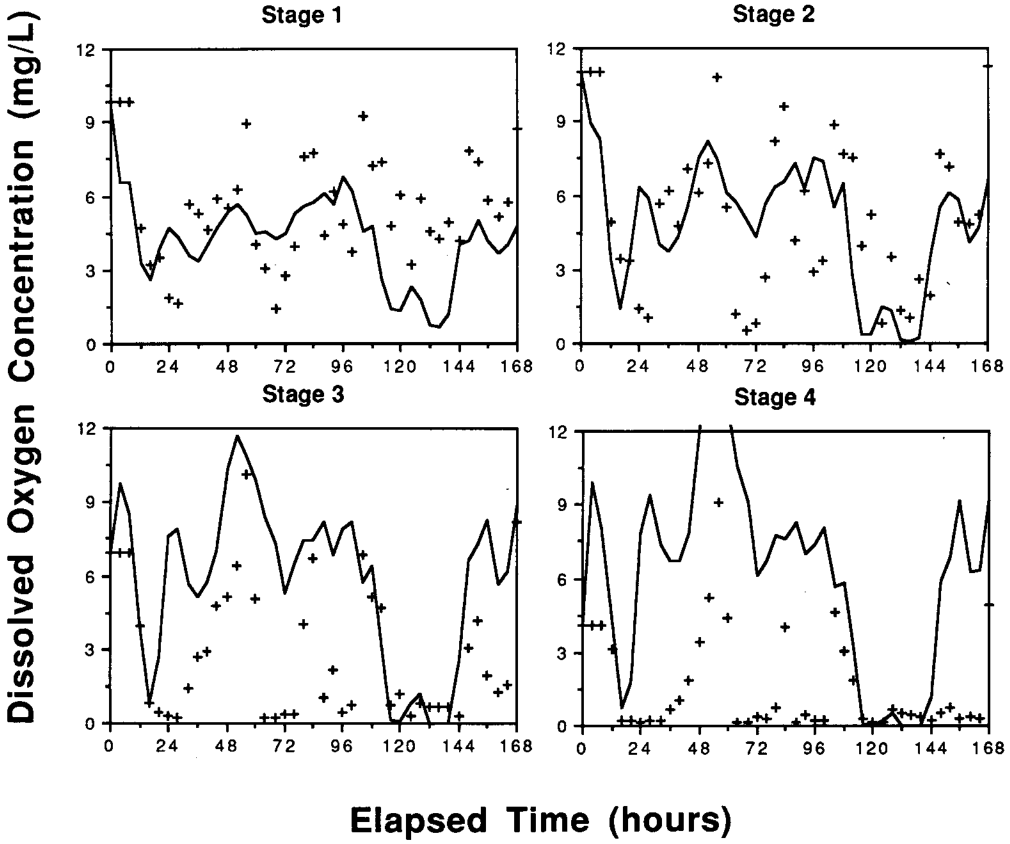

Figure 7—Second process water test simulation: DO.

There was concern that oxygen depletion from absorption to the basin's head space might occur, and that this depletion may influence transfer rates. To prevent oxygen depletion, manhole ventilators were used. The manholes access to the basins were opened and a fresh air cross-flow was established. During testing, head space oxygen purity never fell below 19%. In addition, the turbines were operated at depressed tank DO concentration without gas recirculation to determine the rate of aeration at the tank surface. No measurable change in DO was detected over a 30-minute period.

There was another concern that water quality might affect oxygen transfer performance. According to the manufacturer it was impossible to clean the tanks sufficiently; therefore, test water may not be representatively clean. To evaluate water quality during testing, du Nouy static surface tension measurements were made before, during, and after testing. The measured surface tension of tap water varied, throughout testing, less than ±1 dyne/cm. Alpha factors were also determined for basin clean water as described previously. Oxygen uptake rate was expected to be zero in the basin water and this was confirmed by direct measurement. The α factor associated with basin water was ±5% of unity, which was within the experimental error of the test procedure. It was concluded that the basin was adequately cleaned, and that test water was not contaminated.

The ASCE standard procedure was followed. Experiments were continued to $4/K_La$ units of time and data were analyzed with the nonlinear least-squares procedure. Figure 4 shows the results of the test program reported as standard oxygen transfer rate (SOTR), which is the mass of oxygen transferred per unit time at 20°C, 0 mg/L DO, $\beta = 1.0$, $\alpha = 1.0$, and at atmospheric pressure of 760 mm Hg, and as SAE, which equals SOTR/wire power input. The SAE numbers were calculated based on measured mixer power and the blower power required to produce an equivalent gas flow rate from the plant's recirculation blowers, which was measured previously as 30.9 $m^3/h \cdot kW$ (18.2 scfm/kW). The test blower power was not used because it differed in design and efficiency from the plant blowers.

Stage 3 was tested first. Three repetitions were performed at a gas flow rate of 442 m^3/h (260 scfm). The SOTR values for these three tests were 117.6, 117.8, and 118.3 kg/h, or less than 0.6% difference. The precision of the test was excellent. Based on this reproducibility, only two replicates in the other stages were performed, and the additional test was used to expand the range of gas flow rates. Six probe locations were used at different areas and depths. The probe-to-probe variability was also well within the limits of the standard.

Figure 4 shows optimum gas recirculation rates for each stage. None of the turbines met the specified transfer efficiency of 1.95 kg $O_2/kW \cdot h$ (3.2 lb O_2/hp/hr). Stage 1 was 10% short of optimum gas flow rate. Stages 2 and 3 were 17% short. For the reaeration mode, a higher SAE was anticipated by the manufacturer, Stage 1 was 20%

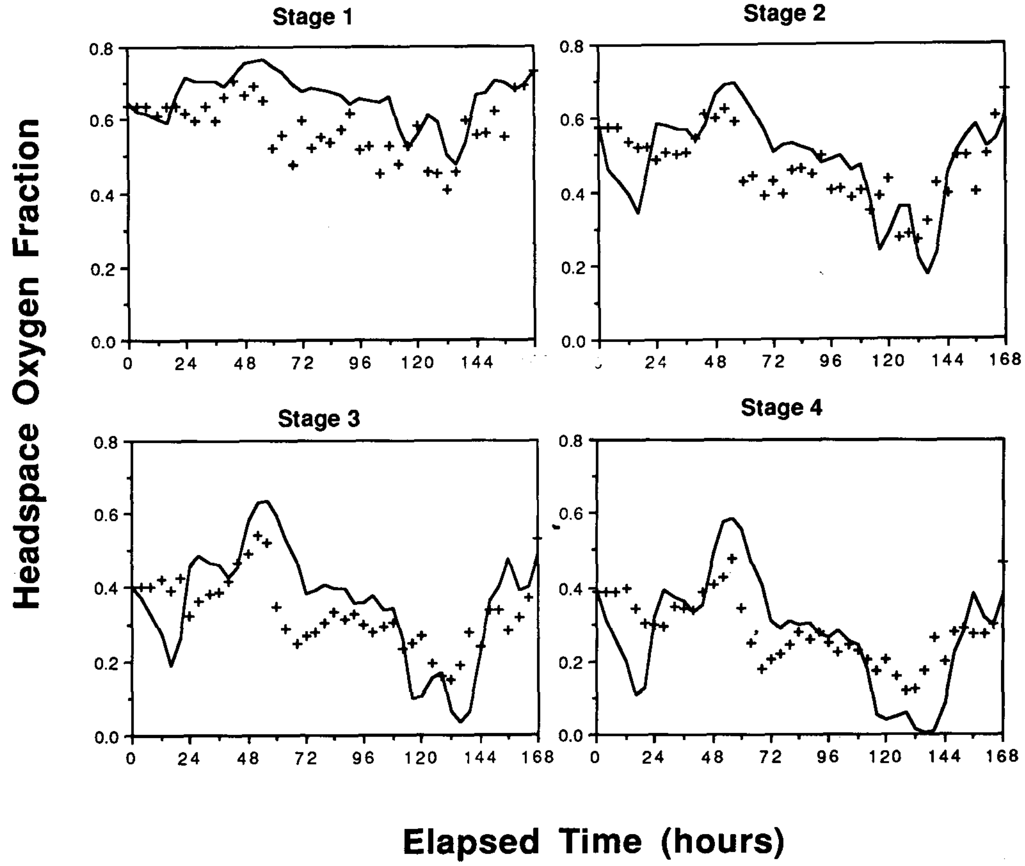

Figure 8—Second process water test simulation: oxygen purity.

short, and Stages 2 and 3 were 25% short. The effective pressure ratio ranged from 1.32 to 1.38, in close agreement with the specifications.

Because the manufacturer did not accept the clean water results as binding, a second process water test was conducted. With the information provided from the clear water testing, the optimum recirculation rates could be used in a second process water test.

Second Process Water Test

The second process water test was conducted in November, 1985. The procedures were very similar to the first test. Six trains were operated. Gas recirculation rates were 1245, 715, 715, and 470 m^3/h in Stages 1 to 4, respectively. This compared to the first process water test recirculation rates of 700, 460, 490, and 490 m^3/h. Alpha factors were determined during the test at 4-hour intervals using the 750-L apparatus. The average α factors for Stages 1 to 4 were 0.63, 0.61, 0.64, and 0.69, respectively. The average α associated with primary effluent was 0.39.

During the test, a large rainstorm occurred affecting plant operation. Also, high-purity oxygen feed rate was upset several times. The cause of those upsets was believed to have been unusual operating conditions needed for the process water test. At one point, the oxygen gas purity in the fourth stage was less than atmospheric purity, decreasing to 11% oxygen. As indicated previously, during low flow in the Sacramento River, the plant diverts effluent to holding basins, and must retreat and discharge the stored effluent during periods of increased river flow rates. During the second process water test it was necessary to divert and retreat effluent.

The manufacturer adjusted Equation 2 based on the specified α and the expected gas purity of 66% and claimed that the plant had passed the performance test. However, the specified transfer rates could not be demonstrated without relying on dubious assumptions regarding alpha factors, steady-state conditions, and gas purity. Steady-state conditions in the aeration basins were never obtained.

Process Modeling

After conducting two full-scale, 7-day process water tests with no definitive conclusions, a dynamic process model was developed to verify or disprove the design oxygen purity profile specified by the manufacturer. The process model was based in part on earlier steady-state models.[5-7] The model is similar in concept to another model.[8] The model, developed here, was written using CSMP III,[9] a simulation program designed to solve systems of linear or nonlinear ordinary differential equations. The model was developed independently of the manufacturer.

Model description. Equations 5 through 21 describe a single stage of a four-stage process. Balances must be written for species in both the liquid and gaseous phase. The model does not include activity coefficients, and is therefore restricted to low ionic strength wastewaters. For the

Table 4—Equations describing a single-stage of a four-stage process.

Species	Equations	
Gas phase		
	$\frac{dCO_2}{dt} = \frac{Q_{Go}CO_{20} - Q_G CO_2}{V_G} - K_L a_{DCD}(DCD_S - DCD_f)\frac{V_L}{V_G MW_{CO_2}}$	(5)
	$\frac{dN_2}{dt} = \frac{Q_{Go}N_{20} - Q_G N_2}{V_G} - K_L a_N(DN_S - DN)\frac{V_L}{V_G MW_{N_2}}$	(6)
	$\frac{dO_2}{dt} = \frac{Q_{Go}O_{20} - Q_G O_2}{V_G} - K_L a_{DO}(C^*_\infty - DO)\frac{V_L}{V_G MW_{O_2}}$	(7)
	$Q_G = K_{Flow}(P_{SP} - P_T)$	
Partial pressures		
	$P_{CO_2} = CO_2 \cdot RT$	(8)
	$P_{O_2} = O_2 \cdot RT$	(9)
	$P_{N_2} = N_2 \cdot RT$	(10)
	$P_T = P_{CO_2} + P_{O_2} + P_{N_2} + P_{H_2O}$	(11)
Liquid phase		
	$\frac{dX}{dt} = \frac{Q_L}{V_L}(X_o - X) + [\mu - K_D]X$	(12)
	$\frac{dDO}{dt} = \frac{Q_L}{V_L}(DO_o - DO) + K_L a_o(C^*_\infty - DO) - \mu X\left(\frac{1-Y}{Y}\right)Y_{O_{21}} - K_D X Y_{O_{22}}$	(13)
	$\frac{dS}{dt} = \frac{Q_L}{V_L}(S_o - S) - \frac{\mu}{Y}X$	(14)
	$\frac{dDN}{dt} = \frac{Q_L}{V_L}(DN_o - DN) + K_L a_{N_2}(DN_S - DN)$	(15)
	$\frac{dDCD}{dt} = \frac{Q_L}{V_L}(DCD_o - DCD) + K_L a_{CO_2}(DCD_S - DCD_f) + \mu X\left(\frac{1-Y}{Y}\right)Y_{CO_{21}} + K_D X Y_{CO_{22}}$	(16)
	$\mu = \frac{\hat{\mu}S}{(K_S + S)} \cdot \frac{DO}{(DO + K_{SDO})}$	(17)
	$DO_S = 5.5555 \cdot 10^4 \frac{MW_{O_2}}{H_{eO_2}} \cdot P_{O_2} \cdot \beta$	(18)
	$DCD_S = 5.5555 \cdot 10^4 \frac{MW_{CO_2}}{H_{eCO_2}} \cdot P_{CO_2} \cdot \beta$	(19)
	$DN_S = 5.5555 \cdot 10^4 \frac{MW_{N_2}}{H_{eN_2}} \cdot P_{N_2} \cdot \beta$	(20)
pH		
	$ALK = [HCO_3] + 2[CO_3^{2-}] + [OH^-] - [H^+] + [NH_3]$	(21)
	$K_W = [OH^-][H^+]$	(22)
	$K_1 = \frac{[H^+][HCO_3^-]}{[H_2CO_3]}$	(23)
	$K_2 = \frac{[H^+][CO_3^{2-}]}{[HCO_3^-]}$	(24)
	$[H^+]^2 + [H^+][ALK - NH_3] - K_W - \left[K_1 + \frac{2K_1K_2}{[H^+]}\right]H_2CO_3 = 0$	(25)

Table 4—(Continued)

Species	Equations	
	$f_{CO_2} = \dfrac{[H_2CO_3]}{[H_2CO_3] + [HCO_3^-] + [CO_3^{2-}]}$	(26)
	$= \dfrac{1}{1 + \dfrac{K_1}{[H^+]} + \dfrac{K_1K_2}{[H^+]^2}}$	(27)
	$DCD_f = DCD f_{CO_2}$	(28)
	$f_{NH_3} = \dfrac{1}{1 + \dfrac{[H^+]K_{NH_3}}{K_W}}$	(29)
	$[NH_3] = [NH_{3T}] \cdot f_{NH_3}$	(30)

Table 5—Nomenclature for model equations.

Symbol	Definition
C^*_{∞}	equilibrium dissolved concentration, g/m^3
CO_2	gas phase carbon dioxide concentration, g moles/m^3
DN	dissolved nitrogen, g/m^3
DCD	dissolved carbon dioxide, including bicarbonate, g/m^3, and carbonate, g/m^3
DO	dissolved oxygen, g/m^3
K_D	decay coefficient, h^{-1}
K_La	volumetric mass transfer coefficient, h^{-1}, includes α
K_S	half saturation coefficient for substrate, g/m^3
K_{SDO}	half saturation coefficient for DO, g/m^3
K_W	ion product in water
K_1	first k_{eq} for carbon dioxide
K_2	second k_{eq} for carbon dioxide
MW	molecular weight, g
N_2	gas phase nitrogen concentration, g moles/m^3
NH_3	undisassociated ammonia concentration at pH, g/m^3
NH_{3T}	total ammonia concentration, g/m^3
O_2	gas phase oxygen concentration, g moles/m^3
Q_L	liquid flow rate per stage, m^3
Q_G	volumetric gas flow rate, m^3
S	substrate, g/m^3
V_L	stage liquid volume, m^3
V_G	stage gas volume, m^3
Y	cell yield, mass X/mass S
$Y_{O_2^1}$	oxygen consumed per unit S consumed
$Y_{O_2^2}$	oxygen consumed per unit X oxidized
$Y_{CO_2^1}$	mass of CO_2 produced per unit mass of S converted
$Y_{CO_2^2}$	mass of CO_2 produced per unit mass of X oxidized
X	cell mass concentration, g/m^3
$\hat{\mu}$	maximum specific growth rate, h^{-1}
o	as subscript, denotes influent value
S	as subscript, denotes saturation concentration at system temperature and partial pressure
f	as subscript, denotes fraction total dissolved carbon dioxide as CO_2 or H_2CO_3
T	as subscript, denotes total pressure
SP	as subscript, denotes set point value of pressure

Sacramento wastewaters (total dissolved solids approximately 500 mg/L), the effects of activity coefficients were assumed to be negligible, and comparisons to a steady-state model[7] validate this assumption. Leakage flows are not shown in this description, but are treated as sinks in the continuity terms. The leakage flow rate was set equal to the measured leakage flow rate. Leakage for the Sacramento case affected the final results by less than 1%. Equation 21 is cubic with respect to $[H^+]$. This equation was reduced to a quadratic by iteratively solving for $[H^+]$ using trial values for the $[H^+]$ in the denominator.

Material balance equations were also written on total ammonia concentration, alkalinity, and inert solids, but are not shown herein. The secondary clarifier was modeled as a zero volume clarifier. Solids thickening was not modeled, as the clarifiers during the periods of the tests were never overloaded. The details of the model are in Tables 4, 5, and 6.

Model results. Figures 5 and 6 show the modeling results and the measured data for the first process water test. The data fit the model well, except in Stage 4, where the model predicts slightly lower oxygen purity. The fit is exceptionally good given that BOD_5 data were determined from analysis of samples that were collected at 24-hour intervals. Undoubtedly, model results would have been better if BOD_5 data were associated with samples collected every 4 hours.

Figures 7 and 8 show the second process water test results. The fit is still good but not as good as in the first

Table 6—Parameter values.

Parameter	Value	Parameter	Value
$\hat{\mu}$	0.20 h^{-1}	$Y_{CO_2^1}$	1.37
Y	0.40	$Y_{CO_2^2}$	1.95
$Y_{O_2^1}$	1.42	K_D	0.004 h^{-1}
$Y_{O_2^2}$	1.42	β	0.99
$Y_{NH_3^1}$	0.039	$K_La_{CO_2}$	0.836 $K_La_{O_2}$
$Y_{NH_3^2}$	0.1239	$K_La_{N_2}$	0.943 $K_La_{O_2}$
K_{SDO}	0.5 g/m^3	θ	1.024
K_s	50 g/m^3		

Table 7—Comparisons of model results and average of measured results.

	DO concentration, mg/L					Oxygen fraction, percent				
	Stage 1	Stage 2	Stage 3	Stage 4	Average	Stage 1	Stage 2	Stage 3	Stage 4	Average
Test 1										
Data	5.1	2.8	1.2	0.6	2.4	66.0	56.6	42.1	32.3	49.3
Model	4.6	1.7	1.1	1.2	2.1	68.6	55.1	41.0	25.8	47.6
Difference	0.5	1.1	0.1	−0.6	0.3	−2.6	1.5	1.1	6.5	1.7
Test 2										
Data	5.5	5.2	2.9	1.6	4.3	58.0	46.9	33.4	28.4	41.6
Model	4.3	4.9	5.8	6.2	5.3	65.8	48.9	36.0	27.5	44.5
Difference	1.2	0.3	−2.9	−4.6	−1.0	−7.8	−2.0	−2.6	0.9	−2.9

test. The second process water test was subjected to two major upsets in high purity oxygen feedrate and large changes in influent concentration and flowrate caused by the rainstorm. In addition, the BOD_5 data from the fifth day of testing were unavailable.

To fit the process data for both tests, a single consistent set of biological parameters was used (Table 6). Only the α factors and process inputs were changed. To improve fit, α factors were adjusted. Initially, the empirically determined α factors were used. The power-weighted α for Test 1 was 0.51. The α factor estimates for Stages 1 to 4 were 0.60, 0.55, 0.40, and 0.40, respectively. In the second test the α factors were reduced to 0.50, 0.50, 0.40, and 0.40 for Stages 1 to 4, respectively, which provided a power-weighted α factor of 0.46.

The α factors estimated in the dynamic modeling compare favorably with those determined experimentally: 0.51 and 0.46 as compared to 0.54 and 0.63 for Test Series 1 and 2, power weighted to match the process tests. Stage 4 contributed more to differences in model and test values. The process test results did not show the increasing trend the pilot scale tests did. This was probably because the power was constant in all pilot tests, and was selected to approximate the average power use in all four stages. In the process tests, the actual power utilization in Stage 1 was 2.1 times that used in Stage 4. Alpha factors can be a function of power density.[9]

Table 7 summarizes average model predictions and averaged experimental results for the two 7-day tests. Arithmetic averages were used for all four stages. Elsewhere, the power weighted average is used. Arithmetic average is useful in comparing goodness of fit. Generally, the agreement is quite good, within 1.7 percentage points for Test 1 and 2.9 percentage points for Test 2, for overall gas purity. The average DO was within 0.3 and 1.0 mg/L for Tests 1 and 2, respectively.

Performance Simulation

The model was used to simulate the warranty conditions. Water quality parameters were set to those shown in Table 2 and α and β factors were set at 0.8 and 0.95, respectively. The results using the manufacturer's suggested gas recirculation rates are shown in Table 8, for both the normal and reaeration modes. Using these recirculation rates, the process can transfer 125 tonnes/d but exceeds the specified power by 240 kW. For the reaeration mode the process also fails, because of insufficient DO in Stage 2.

By adjusting the turbine recirculation rates it is possible to meet the specifications. Table 9 shows that the process just meets the energy requirement while transferring 125 tonnes/d. For the reaeration mode, using all available blowers, the process very nearly meets the specification. The DO concentration in Stage 2 is nearly zero at the warranty condition. To meet the specification, it is necessary to increase the K_La in Stage 2 by increasing turbine horsepower. It was estimated that an 8% increase in turbine horsepower would provide sufficient K_La to meet the specifications.

Conclusions

Compliance testing, which took almost 3.5 years to complete for the Sacramento Regional Treatment Plant, is lengthy and expensive. The greatest difficulty in deter-

Table 8—Model results using manufacturer's gas recirculation rates.

Mode	Flow, m^3/s	Gas recirculation rates, m^3/h	O_2 uptake, tonnes/d	O_2[a] purity, percent	kW	DO,[a] mg/L
Normal	6.0	1275, 764, 713, 713	125	64	1840	5.71
Reaeration	4.0	660, 725, 725, 660	110	61	1656	7.88[b]
Reaeration	6.0	660, 725, 725, 660	156	60	1656	2.37[b]

[a] Power weighted, averaged over all stages.

[b] Insufficient DO, Stage 2.

Table 9—Model results using near-optimal gas recirculation rates.

Mode	Flow, m^3/s	Gas recirculation rates, m^3/h	O_2 uptake, tonnes/d	O_2[a] purity, percent	kW	DO,[a] mg/L
Normal	6.0	1200, 544, 357, 410	125	67	1589	6.0
Reaeration	4.0	1020, 1020, 1020, 663	107	61	1901	12.8
Reaeration	4.3	1020, 1020, 1020, 663	118	61	1901	11.1
Reaeration	4.8	1020, 1020, 1020, 663	130	61	1902	9.5
Reaeration	5.2	1020, 1020, 1020, 663	141	62	1903	7.93
Reaeration	6.0	1020, 1020, 1020, 663	160	62	1904	5.41[b]

[a] Power weighted, averaged over all stages.
[b] Insufficient DO, stage 2.

mining performance compliance is specifying process water and determining α factors. The intent in using the process water test was to warrant several parts of the process beyond the aeration system; however, the shortcoming in the turbine aeration system was obscured by the overwhelming difficulty of measuring ancillary variables, such as α factors.

Alpha factors should have been based on mixed liquor as opposed to the primary effluent. The greatest difficulty in conducting the process water test, particularly the second one, was that the plant was operating very near its maximum transfer rate and efficiency. The shortfall in turbine SOTR and SAE consumed the designer's safety factor; thus, the plant had to operate at optimum conditions to meet the warranty. Optimum conditions are generally unattainable for 7 consecutive days in a large plant like the regional plant.

It is recommended that future performance warranties for high-purity oxygen processes include an ASCE standard clean water test for the aeration system. The mass transfer characteristics of the aeration devices can be accurately established using this procedure. In this work, the ASCE procedure showed replication among tests of ±0.6%.

Planning for the clean water test required approximately 3 months and the test was completed within 1 week. It was possible to clean tanks previously used in the activated sludge process. Less planning would have been required if the ASCE test had been performed before startup, when clean tanks were available.

To ensure other aspects of high-purity oxygen process performance, separate warranties should be written in addition to the ASCE clean water test procedure.

Acknowledgments

Credits. The clean water test was conducted by AERTEC of North Andover, Mass. Larry Effinger performed the alpha factor testing, Ron Linden reviewed the modeling results, and Ed Edmission assisted with the clean water testing.

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