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PARALLEL EVALUATION OF AIR-AND OXYGEN-ACTIVATED SLUDGE

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Contract No. 14-12-150

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Primary differences in performance concerned ammonia nitrogen removals and energy consumption. Differences in sludge production were not significant.

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FOREWORD

The U.S. Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water and spoiled land are tragic testimonies to the deterioration of our natural environment. The complexity of that environment and the interplay of its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution; it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems to prevent, treat, and manage wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, to preserve and treat public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research and provides a most vital communications link between the researcher and the user community.

This report presents design details, operating experiences, and operating and performance data for a parallel operation of an air-activated and an oxygen-activated sludge pilot plant. Consideration of the operational results presented herein is recommended for design engineers, facility planners, and potential municipal users of an oxygen-activated sludge system.

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ABSTRACT

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This report was submitted in fulfillment of Contract No. 14-12-150 by the County Sanitation Districts of Los Angeles County, California, under the sponsorship of the U.S. Environmental Protection Agency. This report covers the operational period of February 9, 1975, to December 23, 1976.

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ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

BOD	-- biochemical oxygen demand
BOD_R	-- biochemical oxygen demand removed
BOD_5	-- 5-day biochemical oxygen demand
COD	-- chemical oxygen demand
COD_R	-- chemical oxygen demand removed
DO	-- dissolved oxygen
DTST	-- deep tank submerged turbine
F/M	-- food-to-microorganism ratio
ISR	-- initial settling rate
JWPCP	-- Joint Water Pollution Control Plant
MCRT	-- mean cell residence time
MLSS	-- mixed liquor suspended solids
MLVSS	-- mixed liquor volatile suspended solids
NTP	-- normal temperature and pressure
RWQCB	-- Regional Water Quality Control Board
SOTR	-- standard oxygen transfer rate
SVI	-- sludge volume index
SWD	-- side water depth
TPVSS	-- total plant volatile suspended solids
VSS	-- volatile suspended solids

SYMBOLS

α, β	-- variables to correlate clean water results to mixed liquor conditions
C	-- system dissolved oxygen concentration
C^*	-- equilibrium dissolved oxygen concentration at zero uptake
dC/dT	-- oxygen transfer rate
θ_c	-- mean cell residence time
k_d	-- microorganism decay coefficient
K_La	-- volumetric mass transfer coeficient
V/Q	-- aeration period
Y	-- growth yield coefficient

SECTION 1

INTRODUCTION

Since the introduction of high-purity, oxygen-activated sludge, a controversy has existed concerning the relative merits of air and oxygen in the activated sludge process, but very few data are available on side-by-side operation of relatively large-scale systems with comparable engineering.

As part of the research effort involved with Federally-mandated secondary treatment at the Joint Water Pollution Control Plant (JWPCP) in Carson, California, the County Sanitation Districts of Los Angeles County constructed two 1900-m^3/day (0.5-mgd) activated sludge demonstration plants. One incorporated the UNOX high purity oxygen process, and one used an air-sparged mechanical aerator. The primary purpose of the study was to obtain data pertinent to the selection and design of an activated sludge system at the JWPCP, but the nature of the research facilities allowed a direct comparison of the two activated sludge processes. The pilot plants were operated on identical feed. Equal engineering care was taken in the design of the aeration systems, and identical clarifiers were used. Unfortunately, the research motivations in establishing the operating parameters for the two plants were different. The oxygen system was operated to refine specified design parameters, while the air system was operated to determine its capabilities and limitations.

The JWPCP is a 15-m^3/sec (350-mgd) primary treatment plant treating a mixture of domestic and industrial wastes. These facilties allowed a good comparison of the two activated sludge alternatives for treating relatively concentrated municipal wastewater.

SECTION 2

CONCLUSIONS

Both activated sludge systems are capable of producing effluents meeting the JWPCP discharge limitations for everything but certain trace metals, which will require source control. But the oxygen system is somewhat more stable and flexible in its operation.

The two systems obtained good removals of soluble organics, and factors affecting solids separation in the final clarifier are most significant in terms of their effects on effluent quality. The most notable detrimental factors encountered in the study were excessive aerator power inputs, which sheared the flocs in both systems, and nitrification-denitrification, which caused the settled sludge from the air system to resuspend.

The major difference between the two systems in terms of pollutant removals concerns ammonia nitrogen. The oxygen system did not nitrify. At the JWPCP, where the ammonia discharge limitation is high enough to impose no constraint, this characteristic is an advantage in that it eliminates rising sludge resulting from nitrification-denitrification.

Claims have been made that oxygen-activated sludge processes produce less sludge than air-activated sludge processes. In this study, a comparison was made based on total plant solids and the difference was found to be insignificant at the 90-percent confidence level. The trend, however, was for the oxygen system to produce more sludge.

Because of modifications to the pilot plant's aeration equipment that were made to prevent floc shear, an energy consumption comparison was considered inappropriate. A paper study indicates that substantial energy savings may be expected with the oxygen system.

SECTION 3

SELECTION AND DESCRIPTION OF THE PILOT PLANTS

AIR-SPARGED TURBINE SYSTEM

The location of the Districts' JWPCP in an urban area placed a definite land constraint on the proposed secondary treatment system for that plant. When preliminary site layouts were made for a conventional activated sludge system with the standard 4.6-m-deep (15-ft-deep) aeration tanks and an optimistic 6-hr aeration period, no excess land was available for waste activated sludge processing. Because of this land constraint, the Sanitation Districts proceeded to evaluate activated sludge systems that could reduce the land area required for secondary treatment. One of those alternatives was the deep tank submerged turbine (DTST) system. The DTST system was selected not only because of the land savings from the deeper tank (7.6 m or 26 ft) but also because the submerged turbine is a more efficient oxygen transfer device than the conventional coarse bubble air diffusers. The land savings from the deeper tank and the possibility of reducing the aeration period made the DTST system a realistic candidate system for secondary treatment at the JWPCP.

The aeration basin for the DTST system (Figure 1) was designed for a 3.5-hr detention time (V/Q) at a design flow of 1900 m^3/day (0.5 mgd). The aeration basin was 6.1 x 6.1 m (20 x 20 ft) with a 7.6-m (25-ft) side water depth (SWD) and 1.5-m (5-ft) freeboard. To insure a complete mix system, 0.51-m (1.7-ft) baffles were provided on each wall running the full tank depth.

The design of the submerged turbine aerator itself was based on an ability to supply sufficient oxygen transfer capability to treat the JWPCP primary effluent in a 2-hr aeration period (V/Q). The turbine aerator had a 45-kW (60-hp) drive unit with a 7.6-m (25-ft) long, 0.25-m (10-in) diameter steel shaft and a 1.5-m (5-ft) diameter impeller. The shaft was supplied in two sections of 6.1 m (20 ft) and 1.5 m (5 ft) to provide the flexibility of evaluating both a 6.1-m (20-ft) and 7.6-m (25-ft) water depth.

Air was introduced into the aeration tank at the perimeter of the mixer/impeller through a sparged ring apparatus. Two 0.28-m^3/sec (10-cfs) air compressors were provided, with one acting as a standby.

HIGH-PURITY OXYGEN SYSTEM

One of the major advantages offered by the pure oxygen biological treatment process is the ability to reduce the period of time required for treatment of a wastewater by increasing the rate at which oxygen can be dissolved into

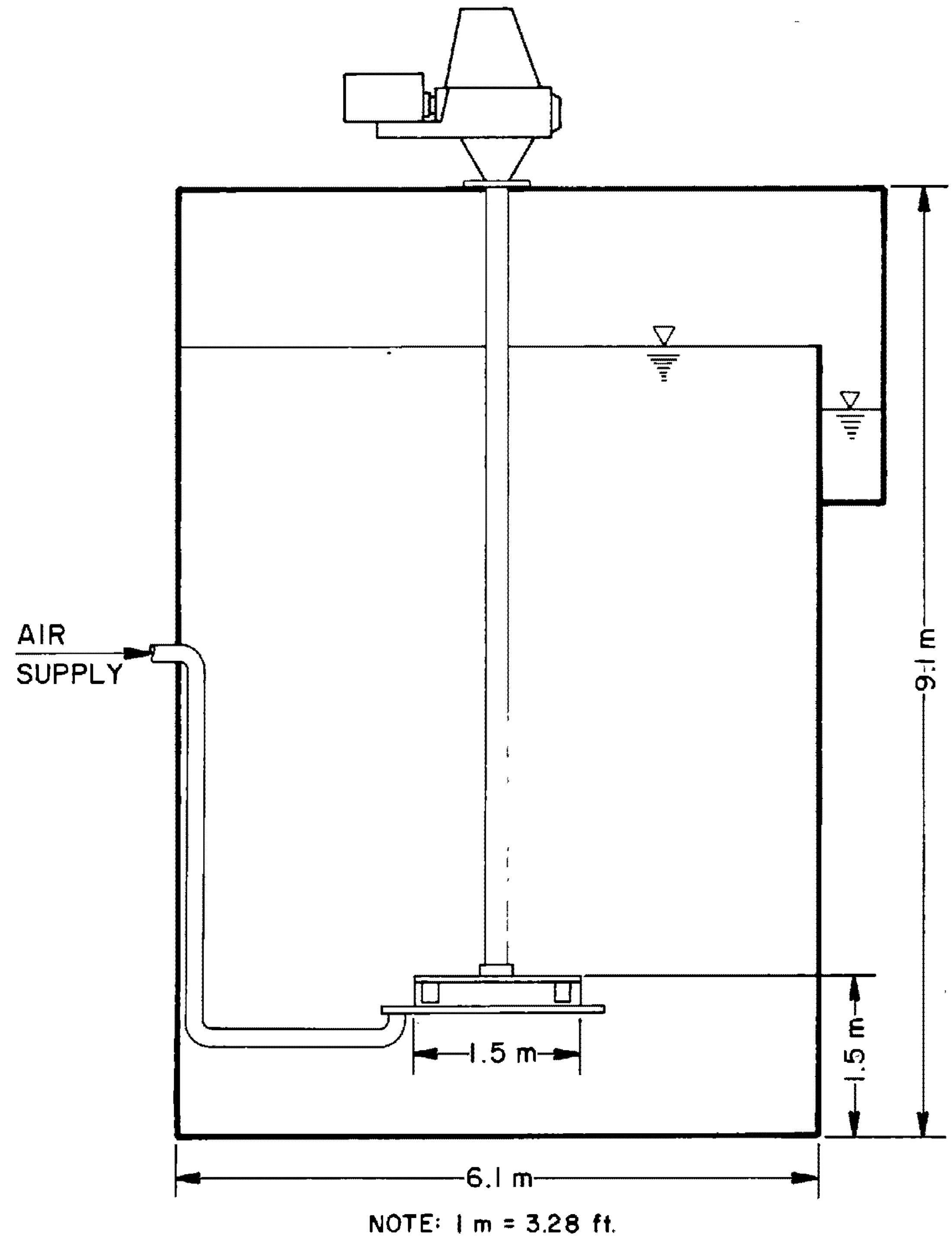

Figure 1. Aeration basin and deep-tank submerged turbine aerator.

the mixed liquor within the biological reactor. The results of preliminary studies using Union Carbide's 0.6-l/sec (10-gpm) mobile pilot plant verified this claim, as acceptable effluent quality was achieved at aeration periods as short as 1.5 hr (V/Q).

Based on this preliminary testing, the oxygen pilot plant was designed for an aeration period of 2.5 hr (V/Q) at the design flow of 1900 m^3/day (0.5 mgd). The biological reactor is 7.3 x 7.3 m (24 x 24 ft) with a 3.7-m (12-ft) SWD. The total height of the basin is 4.6 m (15-ft) (Figure 2). As is typical with the sealed reactor type of pure oxygen system, the reactor was subdivided into four equal-volume, completely mixed chambers with inside dimensions approximating a 3.7-m (12-ft) cube. To insure complete mixing in each of the four reactor stages, there are four anti-swirl baffles per stage located along the diagonals a distance of 1.2 m (4-ft) from the center of the section. These baffles are 0.36-m (1.2-ft) wide and extend the entire depth of the tank. An extension is provided along the bottom 1.8 m (6 ft) of each baffle, which runs toward the tank section center for a total of 0.61-m (2 ft). This modification was included to insure good baffling during operation using surface aerators, if so desired.

As a result of competitive bidding, Union Carbide Corporation was awarded a contract for the construction of the pure oxygen biological reactor, which was to be built into the existing pilot plant influent pumping station and final clarifier system. The reactor was designed to incorporate a submerged turbine/gas recirculation compressor arrangement for oxygen dissolution in each reactor stage. The mixers in stages 1 and 2 were driven by 3.7-kW (5-hp) motors, while those in stages 3 and 4 were driven by 2.2-kW (3-hp) motors (Figure 2).

Having been introduced into the gas space above the liquid level in stage 1 of the reactor, the oxygen was withdrawn from the gas space above the stage 1 mixed liquor level by a compressor and pumped through the center of the 0.15-m (6-in.) diameter turbine shaft. The gas exited the shaft through a rotating sparger located approximately 0.3 m (1 ft) from the bottom of the reactor at the base of the shaft. Four rectangular turbine blades were located about 0.3 m (1-ft) above the rotating sparger, which, when operated at their normal speeds (130 rpm in stages 1 and 2, 82 rpm in stages 3 and 4), maintained a completely mixed regime while dissolving sufficient amounts of oxygen to meet the biological demand. Oxygen which did not go into solution and carbon dioxide coming out of solution as a by-product of the biological reaction in the first stage passed through an opening in the gas space into the second stage where it was introduced into the mixed liquor by the same compressor/turbine arrangement as the first stage. In like manner, the gas proceeds through the third and fourth stages of the reactor, with the unused oxygen and other gases being ultimately passed through a vent in the fourth stage to the atmosphere.

The dissolved oxygen concentration in each stage was controlled by varying the recirculated gas flow from the compressor to the sparger at the base of the turbine shaft. This was accomplished by means of a 50-mm (2-in.) bypass valve located between the compressor discharge and a rotary joint gas inlet

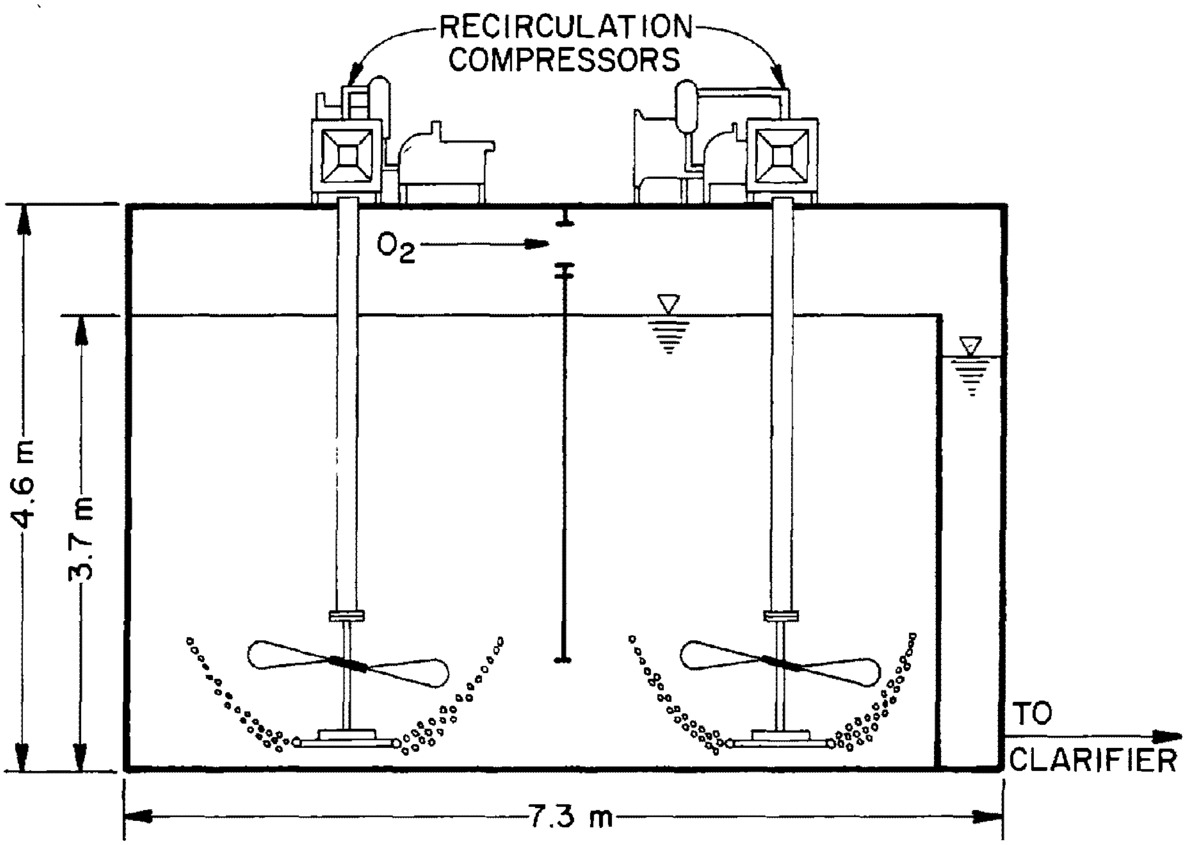

Figure 2. Oxygen system reactor.

at the top of the turbine. By opening this valve, a portion of the recirculation gas can be bypassed back into the gas space, thus reducing the volume of gas which is introduced into the liquid through the turbine for dissolution.

The oxygen supply system is controlled through a pressure transmitter-set point controller arrangement in which the flow of oxygen from the storage facility to its point of introduction at the first stage is automatically controlled in order to maintain a constant pressure in the gas space above the first stage. When the pressure controller indicates a pressure that is below the pressure set point, an increased signal is sent to a control valve, which increases the pure oxygen gas flow into the reactor and, hence, the pressure in the gas space over the first stage. Likewise, if the gas space pressure exceeds the set point value, the gas flow is reduced until the pressure approaches the desired value.

The flow of gas through the reactor is monitored at all times so that the amount of oxygen utilized during the treatment process can be determined. Since the reactor is sealed, the monitoring of oxygen utilization is accomplished by simply measuring the mass flow rate of pure oxygen into the reactor and subtracting the mass flow rate of oxygen exiting the reactor via the vent stack in the fourth stage. In the latter case, it is necesary to measure and record the oxygen composition of the vented gas continuously, since a significant portion is composed of gaseous byproducts of the chemical and biological reactions that take place while the wastewater is under aeration.

FINAL CLARIFIERS

Three final sedimentation tanks were designed for the project using the Districts' basic criteria for rectangular final sedimentation tanks. Two of the tanks were of the same size to allow evaluation of both the submerged turbine system and the high purity oxygen system at the same overflow rate of 28.5-m^3/m^2/day (700-gpd/ft^2) at the design flow 1900-m^3/day (0.5 mgd). The third tank was designed for an overflow rate of 18.3 m^3/m^2/day (450 gpd/ft^2) at the 1900-m^3/day (0.5-mgd) flow. It was used to evaluate lower overflow rates in either system and to provide the flexibility required to evaluate lower aeration times and, hence, flows of greater than 1900-m^3day (0.5 mgd) in either pilot plant.

The two final sedimentation tanks designed for 28.5-m^3/m^2/day (700 gpd/ft^2) were 3-m (10-ft) deep, 3-m (10-ft) wide, and 22-m (72-ft) long. These tanks have a 2-hr hydraulic detention time and a flowthrough velocity of 3.2 mm/sec (0.6 ft/min) at the 1900-m^3/day (0.5-mgd) flow and 30-percent recycle. The third final sedimentation tank had the same width and depth as the two 22-m (71-ft) tanks, but it was 34-m (111-ft) long. The hydraulic detention at 1900-m^3/day (0.5-mgd) flow and 30-percent recycle was 3 hr, and because it had the same cross sectional area as the shorter sedimentation tank, the flowthrough velocity was the same. The same weir length was provided on all three sedimentation tanks, so that at the design flow, the weir loading was 62.1-m^3/m/day (5000-gpd/ft^2).

The design criteria used for the biological reactors and the associated final sedimentation tanks have been summarized in Table 1.

TABLE 1. DESIGN CRITERIA FOR PILOT PLANTS

Item	Air System	Oxygen System
Biological Reactors:		
Average flow, m^3/day (mgd)	1900 (0.5)	1900 (0.5)
Length, m (ft)	6.1 (20)	7.3 (24)
Width, m (ft)	6.1 (20)	7.3 (24)
Average water depth, m (ft)	7.6 (25)	3.7 (12)
No. of stages	1	4
Detention time (V/Q), hr	3.5	2.5
Oxygen Storage Tank:		
Number	--	1
Volume, m^3 (ft^3) NTP	--	9900 (350,000)
Capacity, m^3/hr (ft^3/hr)	--	140 (4940)
	Standard	Large
Final Clarifiers:		
Number	2	1
Length, m (ft)	22 (72)	34 (111)
Width, m (ft)	3.0(10)	3.0 (10)
Average water depth, m (ft)	3.0(10)	3.0 (10)
Overflow rate, m^3/m^2/day (gpd/ft^2)	28.5(700)	18.3 (450)
Detention time (Q + 1/3 return), hr	2.0	3.0
Weir loading rate, m^3/m/day (gpd/ft)	62.1 (5000)	62.1 (5000)
Flowthrough velocity (Q + 1/3 return), mm/sec (ft/min)	3.2 (0.6)	3.2 (0.6)

SECTION 4

OPERATION OF THE PILOT PLANTS

STARTUP

Air Sparged Turbine Pilot Plant

Upon completion of the clear water testing of the DTST aerator in December 1974, the DTST system was started up in January 1975. The pilot plant was seeded with waste activated sludge from the Pomona Water Reclamation Plant. From the middle of January until mid-February, the flow to the unit was gradually increased from 380-to 1100-m^3/day (0.1-to 0.3-mgd). However, during this period, the effluent was characterized by cloudiness and the biology was marked by an apparent dispersed floc. A meeting with the mixer manufacturer's representatives was called in mid-February. The discussions indicated that the probable cause for high effluent turbidity and dispersed floc was shearing of the floc. To alleviate this problem the manufacturer agreed to decrease the energy input to the basin by reducing the aerator speed from 54 to 46-rpm. The mixer horsepower was thereby reduced 37-percent. Once the mixer speed was reduced, the improvement in effluent quality was almost immediate. Within a few days, the cloudiness in the effluent disappeared and a good biological floc appeared.

UNOX Pilot Plant

The oxygen biological treatment pilot plant was started up on June 27, 1975, by drawing air into the reactor through the recirculation gas compressors with no seed being added. The system responded very quickly, and by July 15, 1975, what appeared to be a good, stable sludge had been achieved. A series of mechanical difficulties was encountered at this time that hindered the normal progression of operation toward a steady-state condition. However, after almost 45 days of operation, during which the unit had been seeded, it became apparent that continued poor effluent quality (high turbidity and suspended solids) was the result of causes other than these mechanical startup difficulties.

During this period, the system was operated over various hydraulic and organic loading rates and investigations were made as to possible toxic compounds in the primary effluent. However, toxicity was soon dismissed as a possible cause of poor effluent quality, not only by an examination of primary effluent trace constituent concentrations, but also by the fact that the DTST system was being operated concurrently without showing any signs of toxic effects.

Through further investigation, other possible causes (such as low pH and floc shear through excessive turbine blade tip speeds) were eliminated. The major factor was finally traced to an energy intensity problem related to oversized gas recirculation compressors and resulting floc shear due to the flooding of the spargers by excessive pumping rates. An expedient solution was achieved in early September 1975 by drastically reducing the flow of recirculated gas, the result of which was significant improvement in effluent quality in general and a decrease in turbidity in particular. The improvement was still not to the level that had been achieved in 1973 during the operation of Union Carbide's 0.6-l/sec (10-gpm) mobile pilot plant, but the effluent being produced was within the State and Federal discharge requirements.

As outlined earlier, the pure oxygen pilot plant was originally designed with provisions made for conversion from submerged turbines to surface aerators at a later date, if so desired. However, with the accelerated State construction grants program and the ensuing decision to design a full-scale oxygen surface aeration system at the JWPCP, immediate steps were taken to convert the pilot plant to a surface aeration system.

On September 25, 1975, the pilot plant was taken out of service following a short period of good operation under diurnal flow conditions. On October 3, 1975, the installation of the surface aeration equipment was completed and the system was restarted. The influent flow was gradually increased to 1500 m^3/day (0.4 mgd), and beginning on October 23, 1975, the first period of good steady-state operation was obtained and was subsequently sustained for a 3-wk period. Following this period, it was intended that the influent feed flow be changed to simulate the JWPCP diurnal flow pattern but difficulties relating to the operation of the system using surface aerators prevented this progression.

Soon after the system was restarted with the surface aerators installed, a great deal of gas was observed escaping above the clarifier inlet diffusers. In addition, the oxygen utilization data gathered during the surface aerator operation was not at all in agreement with similar data gathered both during the earlier operation using submerged turbines and during the 1973 operation of the 0.6-l/sec (10-gpm) mobile pilot plant. It was assumed, therefore, that gas from the fourth stage of the reactor was somehow being trapped within the mixed liquor and was subsequently being purged as the liquid entered the final clarifier. It was theorized that the only way in which such large volumes of gas could be conveyed out of the reactor and into the mixed liquor piping would be the result of the aerator umbrella creating excessive turbulence in the trough downstream of the overflow weir in the fourth stage of the reactor as illustrated in Figure 3. With such unmeasured quantities of gas escaping, it was impossible to accurately measure the critical parameter of oxygen utilization.

In mid-December 1975, a baffle was installed in front of the overflow weir in the fourth stage of the pilot reactor by representatives of Union Carbide. The purpose of the baffle was to prevent the aerator umbrella from extending into the trough downstream of the weir. This baffle, however, was not sufficient as the gas leakage was reduced but not eliminated entirely. It became clear that in order to completely correct the problem, the pilot plant

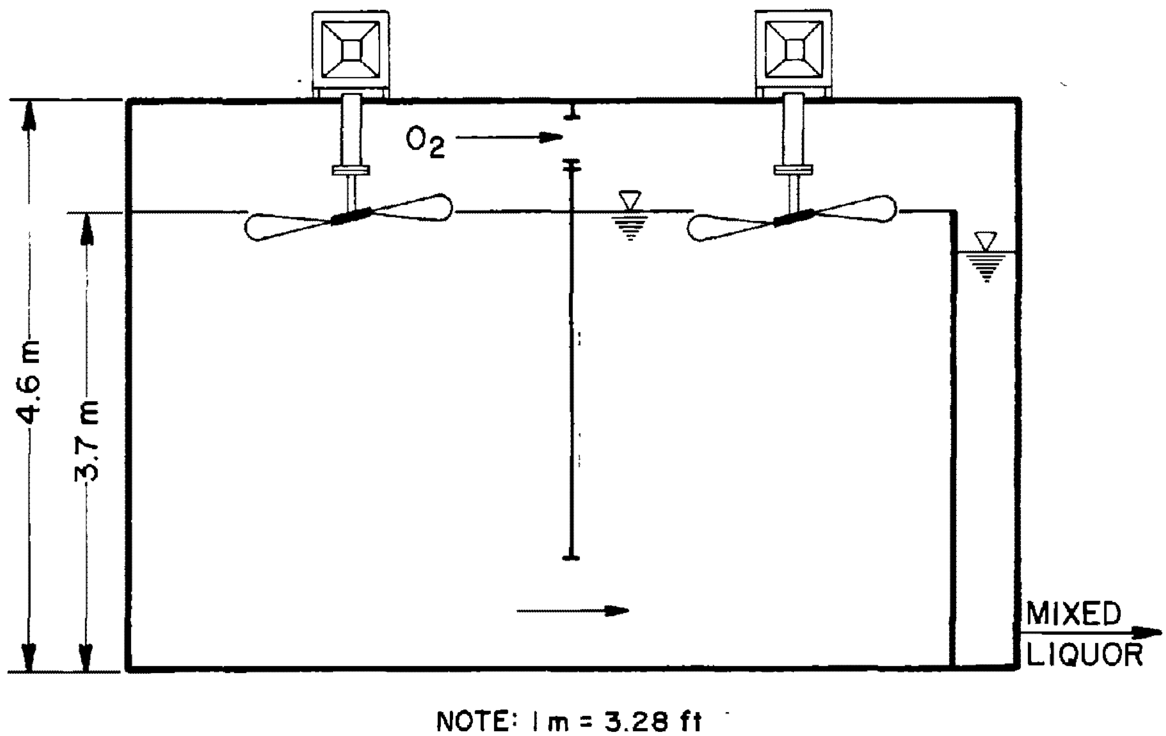

Figure 3. UNOX reactor with surface aerators.

would have to be taken out of service and a much larger baffle installed. Since a short timetable was available for completion of the first phase of the full-scale oxygen secondary treatment design, it was necessary that the research studies of design criteria and the actual design of the secondary system be conducted simultaneously. Because of this, it was decided that operation of the pilot plant be continued despite the difficulty in assessing oxygen utilization.

The next 2-mo were spent developing methods of improving sludge settleability so that critical design parameters related to the secondary clarifiers could be evaluated. By mid-March 1976, good, steady operation had been established and a 5-mo study of clarifier performance was begun. By mid-April, however, it became apparent that even though steady-state operation was being maintained, the effluent turbidity was still not equal to that which has been obtained earlier during the mobile pilot plant studies. Microscopic studies of the secondary effluent led to the conclusion that the floc was being sheared by the aerators to a certain extent, which was the cause of the cloudiness in the effluent. On April 29, 1976, the mixer speed in the fourth stage of the reactor was reduced from 68 rpm to 45 rpm, which represents a reduction in power of about 33 percent.

The effect of the power reduction in the fourth stage of the reactor was two-fold. First, the effluent turbidity was improved as expected following the change. Second, the gas leakage into the clarifier was further reduced but not eliminated. Following the extended steady-state operating period and clarifier evaluation, the pilot plant was taken out of service and a larger, more permanent baffle was installed in place of the one installed earlier. On September 13, 1976, the pilot plant was re-seeded and since that time the system has performed very well. There is no longer any gas leakage into the clarifiers, and useful oxygen utilization data have become available.

PILOT PLANT OPERATIONAL PHASES

Air Sparged Turbine Pilot Plant

As previously mentioned in the startup subsection, initial startup operational problems were encountered from the high energy input to the aeration basin, which were manifested in shearing of the floc. After these startup problems were resolved in mid-February through slowing down of the aerator's speed, the pilot plant started its first steady-state phase in February 1975. The time period of February 1975 through March 1976 has been divided into nine steady-state operational phases. The basic criteria used in defining steady-state operational phase were the mean cell residence time (MCRT or θ_c) and aeration period (V/Q). These two major operational parameters or independent variables were held constant for a given mode of operation. The resulting operational data for the nine phases are summarized in Table 2.

The pilot plant operational phases can be further divided into two areas. Phases I through VI were conducted to determine the operational limitations of the DTST system and to verify the organic and trace constituent removals that the diffused air activated sludge pilot plant achieved during a previous study. Although Phases VII through IX do not show much variation between the basic operational parameters of MCRT and aeration period, extensive testing of the final clarifiers was conducted during these phases. During Phases VII through IX, a secondary operational parameter, recycle rate, was varied to determine its effect on the solids inventory, clarifier hydraulics, and loading rates. Also, the DTST operation for Phases VIII and IX was conducted to provide parallel operation data for comparison with the oxygen pilot plant. Although Phases VIII and IX do not correspond to a specific phase of operation for the oxygen system, they do represent parallel operational periods and, for the most part, all of the pilot plant data can be used to compare the two types of systems based on similar operational conditions.

Phase I represents the first steady-state operational period of the DTST pilot plant. During this phase, the pilot plant was operated at a 5.6-hr aeration period and a 6.8-day MCRT was maintained. The 7-day MCRT was maintained to keep a high level of solids within the system. These solids were maintained to ease the transition to the shorter aeration periods and higher loadings for which the system was designed. Under these operational conditions, partial nitrification was achieved. The partial nitrification and the long detention time in the final clarifier resulted in denitrification and, hence, rising sludge in the final clarifier. To alleviate the rising

TABLE 2. SUMMARY OF OPERATIONAL PARAMETERS -- AIR-SPARGED TURBINE SYSTEM

PARAMETER	PHASE I	II	III	IV	V	VI	VII	VIII	IX
DATES									
Start	2/9/75	3/9/75	4/6/75	5/11/75	7/20/75	9/28/75	10/26/75	11/27/75	3/4/76
End	3/1/75	3/29/75	5/3/75	6/21/75	8/30/75	10/25/75	11/20/75	12/25/75	3/25/76
Duration, days	21	21	29	42	42	28	26	29	22
Flow Pattern	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Steady
REACTOR									
Influent Flow, m^3/day (mgd)	1200 (0 32)	1700 (0 45)	1700 (0 45)	1900 (0 50)	1700 (0 45)	1500 (0 40)	1500 (0 40)	1500 (0 40)	1300 (0 34)
Recycle, %	90	65	45	40	44	29	38	50	47
Hydraulic Detention Time									
V/Q, hr	5 6	4 0	4 0	3 5	4 0	4 5	4 5	4 5	5 3
V/(Q+R), hr	2 9	2 4	2 8	2.5	2 8	3 5	3 3	3 0	3 6
MLSS, mg/l	3100	3400	2600	4000	2300	3300	3300	3600	2900
Volatility, %	72	73	74	73	73	70	71	70	70
Mean Cell Residence Time									
Reactor Solids, days	5 1	3 7	2 2	3 7	1 8	3 0	3 2	3 4	3 6
Total System Solids, days	6 8	5 4	3 3	5 5	2 8	4.3	4 3	4 5	5 9
Organic Loading Rate									
BOD_R/MLVSS, kg/kg/day	0 34	0 38	0 49	0 30	0 70	0 49	0 44	0 44	0 45
BOD_R/TPVSS, kg/kg/day	0 26	0 27	0 33	0 23	0 47	0 33	0 30	0 30	0 29
COD_R/MLVSS, kg/kg/day	0 80	1 07	1 30	0 90	1 61	1 16	1 00	1.00	1 10
COD_R/TPVSS, kg/kg/day	0 60	0 74	0 87	0 60	1 06	0 82	0 75	0 75	0 68
BOD_A, kg/m^3/day (lb/ft^3/day)	0 75 (12 0)	1 00 (16 0)	1 03 (16 5)	1 15 (18 4)	1 34 (21 5)	1 24 (19 9)	1 12 (17 9)	1 20 (19 2)	0 97 (15 5)
Sludge Production									
VSS/BOD_R, kg/kg	0 51	0 64	0 79	0 73	0 70	0 56	0 63	0 63	0 60
VSS/COD_R, kg/kg	0 22	0 27	0 34	0 30	0 35	0 26	0 31	0 30	0 27
CLARIFIER									
Overflow Rate, m^3/m^2/day (gpd/ft^2)	18 3 (450)	21 3 (523)	16 9 (415)	18 3 (450)	16 1 (395)	14 7 (361)	14 7 (361)	14 7 (361)	19 4 (476)
Detention Time									
V/Q, hr	4 0	2 8	4 3	4 1	4 5	5 0	5 0	5 0	3 6
V/(Q+R), hr	2 1	1.7	3 0	2 9	3 1	3 9	3 6	3 3	2 5
Solids Loading Rate, kg/m^3/day (lb/ft^3/day)	107 (1714)	117 (1874)	63 (1009)	103 (1650)	54 (865)	63 (1009)	68 (1089)	83 (1329)	83 (1329)
Return Sludge Concentration, %	0 7	0 9	0 9	0 9	0 9	1 2	1 1	1 1	0 9
SVI, ml/g	252	183	163	165	227	200	160	173	146

sludge problem, the sludge was removed as rapidly as possible from the final clarifier as indicated by the 90-percent recycle rate.

As the system showed signs of stabilizing, the aeration time was decreased to 4.0-hr and the MCRT was reduced to 5.4 days. At these conditions, the DTST system was able to maintain good organic removals and effluent clarity, but rising sludge was still a problem, which again resulted in an inordinate amount of solids being carried over the weir into the effluent.

During Phase III operation, the aeration period was maintained at 4.0 hr, but the MCRT was lowered from 5.4 to 4.0 days. Under these conditions the organic removals remained good. The problem of solids carry over in the final effluent was alleviated by switching to the longer final clarifier as indicated by the lower over flow rate of 16.8 m^3/m^2/day (412 gpd/ft^2)

The aeration period was lowered to 3.5 hr in Phase IV, and to maintain reasonable loading rates on the system at this short aeration period, the plant solids were increased by increasing the MCRT to 5.6 days. Good treatability was observed under these operational conditions.

Phase V operation constituted the highest sustained loading period of the study for the DTST pilot plant. Although the aeration time was increased slightly to 4.0 hr, the MCRT was reduced to 2.8 days. Even though the DTST was able to treat the wastewater under these conditions, the pilot plant was extremely sensitive to operate. This was reflected by a 2-wk period within this phase when the effluent suspended solids averaged 30 mg/l. The pilot plant, however, soon reached an overloaded condition after this short period of good operation, and the effluent quality started to decline.

The aeration period was increased to 4.5-hr, and the MCRT was increased to a more manageable 4.3-days in Phase VI. The DTST system responded to these operational changes, and stable operation of the pilot plant resumed.

Phases VII and VIII were a continuation of Phase VI with the aeration period and MCRT remaining the same for all three phases. However, the 30-percent recycle rate in Phase VI was increased to 40 percent in Phase VII and 50-percent in Phase VIII. During these phases, the effect of the sedimentation tank hydraulics, overflow rate, and solids loading rate on the thickening of the return sludge was studied. Also, the effect of the recycle rate on the mass flow back to the reactor was studied.

The aeration period was further increased to 5.3 hr in Phase IV while the MCRT was increased to 5.9 days. This operational mode was run to see if the DTST system could operate under conventional conditions and not have the nitrification-denitrification problems that were associated with Phase I.

UNOX Pilot Plant

Simply stated, the major objectives of the high purity oxygen pilot plant studies were twofold: first, to gather information that would be pertinent to the full-scale treatment plant design effort that was being conducted concurrently and, second, to develop operational techniques which could simplify the startup and operation of this full-scale system. The 0.6-l/sec (10-gpm) mobile pilot plant had provided treatability information and data to allow some equipment sizing, but certian key design questions were left unanswered at the completion of the mobile pilot plant testing. First, the clarifier used during the preliminary studies was an unconventional circular model that provided low overflow rates and a great deal of sludge storage capacity. Since the full-scale system would be operated using rectangular clarifiers that were smaller in relation to the biological reactor than had been the case during the preliminary studies, it was imperative that the performance of rectangular clarifiers be evaluated. This evaluation is critical since the operation of a high purity oxygen system is generally limited by the ability of the secondary clarifier to store and convey sludge solids.

The second key question to be addressed by the 1900-m^3/day (0.5-mgd) plant operation concerned the system oxygen requirements, particularly the daily fluctuation in oxygen demand, which is a result of the diurnal variation in flow and organic loading at the JWPCP. Information in this regard would have a direct bearing on the selection of equipment for the cryogenic oxygen generating system that is being provided to supply oxygen to the biological treatment system.

The priorities of the 1900-m^3/day (0.5-mgd) pilot project following the July 1975 startup were to stabilize the system at design conditions as quickly as possible and to collect data relative to the required design information. Beyond this, information regarding system limitations and overall operation would be documented. This phase of operation would require the more rigorous approach to pilot operation of biological treatment systems wherein the system performance would be evaluated over an entire range of organic loading rates and aeration periods.

As a result of the startup difficulties outlined earlier in this report, acceptable operation of the pilot plant could not be achieved before late September 1975. Only 4 days of good operation (Phase I) were recorded before the pilot plant was taken out of service on September 25, 1975, for the installation of surface aerators. From this point until mid-October 1976, the operation of the pilot plant has been divided into eight periods, which are representative of good steady operating periods and/or periods during which specific objectives were being met. Operational parameters are summarized in Table 3.

Phase I, though it includes only 4 days of testing, is significant in that it represents the first successful pilot operation in which the system was operated under a simulated diurnal plant flow condition.

Phase II represents the first period of good operation following the installation of surface aerators in the biological reactor. During this period, attempts were made to stabilize the operation at the 7-day design MCRT in order to begin the evaluation of the rectangular clarifier as well as to further establish the organic removal and oxygen demand relationships. It was during this period, however, that the gas "boiling" problem outlined earlier was first discovered. By late October 1975, the difficulty became

TABLE 3. SUMMARY OF OPERATIONAL PARAMETERS -- OXYGEN SYSTEM

PARAMETER	PHASE										
	I	II	III	IV	V	VI	VII	VIII	IX	X	XI
DATES											
Start	9/22/75	10/27/75	12/1/75	2/1/76	2/18/76	3/31/76	6/21/76	9/30/76	10/28/76	11/9/75	12/10/76
End	9/25/75	11/10/75	12/30/75	2/17/76	2/29/76	5/20/76	9/14/76	10/13/76	11/7/76	11/24/76	12/23/76
Duration, days	4	15	30	17	12	51	85	14	11	16	14
Flow Pattern	Diurnal	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Diurnal	Diurnal	Diurnal
REACTOR											
Influent Flow, m^3/day (mgd)	1900 (0.51)	1500 (0.40)	1400 (0.37)	1700 (0.45)	1900 (0.51)	1900 (0.51)	1800 (0.48)	1900 (0.51)	1900 (0.51)	1600 (0.43)	1600 (0.43)
Recycle, %	40	40	44	44	42	40	38	40	39	47	39
Hydraulic Detention Time											
V/Q, hr	2.5	3.1	3.4	2.8	2.5	2.5	2.6	2.5	2.5	3.1	3.0
V/(Q+R), hr	1.8	2.2	2.3	1.9	1.6	1.8	1.9	1.8	1.8	2.1	2.2
MLSS, mg/l	3800	2800	4200	4600	3300	3900	4100	4420	3700	3990	3840
Volatility, %	75	73	74	72	75	74	77	75	70	70	77
Mean Cell Residence Time											
Reactor Solids, days	1.8	2.5	3.4	1.9	1.7	1.9	2.7	2.1	2.0	3.0	2.8
Total System Solids, days	3.4	5.9	6.8	5.6	3.4	4.4	4.8	3.8	4.2	6.6	5.4
Organic Loading Rate											
BOD_R/MLVSS, kg/kg/day	0.70	0.74	0.52	0.60	0.83	0.69	0.57	0.48	0.67	0.55	0.51
BOD_R/TPVSS, kg/kg/day	0.31	0.31	0.26	0.20	0.42	0.29	0.33	0.27	0.32	0.24	0.27
COD_R/MLVSS, kg/kg/day	1.67	1.52	1.14	1.31	1.61	1.54	1.15	0.95	1.46	1.07	1.05
COD_R/TPVSS, kg/kg/day	0.89	0.64	0.56	0.45	0.81	0.64	0.66	0.54	0.69	0.47	0.55
BOD_A, kg/m^3/day (lb/ft^3/day)	2.15 (34.4)	1.73 (27.7)	1.62 (25.9)	2.03 (32.5)	2.05 (32.8)	2.00 (32.0)	1.76 (28.2)	1.63 (26.1)	1.94 (31.0)	1.54 (24.6)	1.44 (23.0)
Oxygen Utilization											
O_2/BOD_R, kg/kg	1.36	--	--	--	--	--	--	1.52	1.24	1.48	1.49
O_2/COD_R, kg/kg	0.71	--	--	--	--	--	--	0.81	0.69	0.71	0.70
Sludge Production											
VSS/BOD_R, kg/kg	0.97	0.60	0.64	0.63	0.78	0.80	0.69	0.84	0.98	0.74	0.66
VSS/COD_R, kg/kg	0.48	0.29	0.28	0.29	0.40	0.36	0.33	0.42	0.38	0.38	0.37
CLARIFIER											
Overflow Rate, m^3/m^2/day (gpd/ft^2)	18.7 (459)	23.2 (570)	21.2 (521)	25.4 (625)	28.4 (698)	27.9 (686)	27.5 (676)	18.1 (445)	28.4 (698)	23.3 (573)	23.2 (570)
Detention Time											
V/Q, hr	3.7	3.0	3.3	2.8	2.4	2.5	2.5	3.8	2.5	2.9	2.9
V/(Q+R), hr	2.8	2.2	2.3	1.9	1.7	1.8	1.8	2.7	1.8	2.0	2.1
Weir Loading Rate, m^3/m/day	79.1	62.6	52.2	68.9	77.0	101.2	99.4	101.5	102.3	84.2	85.8
(ft^3/ft/day)	(852)	(674)	(562)	(741)	(829)	(1089)	(1070)	(1092)	(1101)	(906)	(923)
Solids Loading Rate, kg/m^3/day	98	90	127	168	134	152	141	113	147	141	126
(lb/ft^3/day)	(1568)	(1440)	(2032)	(2688)	(2144)	(2432)	(2256)	(1808)	(2352)	(2256)	(2016)
Return Sludge Concentration, %	1.05	1.06	1.40	1.54	1.18	1.36	1.22	1.34	0.88	0.99	0.94
SVI, ml/g	78	153	99	65	83	77	83	113	124	114	101

clearly defined and a decision was made to forego the oxygen utilization investigations until a later date so that the evaluation of the final clarifier could proceed.

The turbulence created at the clarifier inlet by the escaping gas resulted in an unusual amount of solids being lost through the clarifier skimming system. Because of the difficulty in both measuring and controlling this solids loss, the actual phase average cell MCRT was less than the desired 7-day level. Attempts at controlling this solids loss to sustain good operation at the desired MCRT ultimately resulted in the loss of steady-state conditions and an end to this phase of the pilot operation.

Because of the construction at the JWPCP, it became necessary to relocate the pump suction lines of the pilot plant influent pump station. As a result of this change, it was not possible to operate the pilot plant at the design flow rate (1900 m^3/day or 0.5 mgd) during most of December 1975. Though, by strict definition, a steady-state condition was never achieved during this period, Phase III of the pilot plant study represents a period of good stable operation under adverse conditions. During this period, attempts were made to improve sludge settleability. Moreover, the first attempt was made toward correcting the gas leaking problem outlined earlier through the addition of a baffle by representatives of Union Carbide.

During January and the early part of February 1976, several attempts were made to stabilize the operation at both the design flow (1900 m^3/day or 0.5 mgd) and cell MCRT (7 days). While stable operation was achieved under these conditions, it became apparent by mid-February 1976 that the 7-day MCRT residence time could not be maintained without severely stressing the final clarifier. While good organic removal and sludge settleability were evident, the sludge blanket levels that were necessary to sustain this mode of operation resulted in poor effluent quality and, hence, unacceptable operation. Phase IV from Table II summarizes this period of operation.

Following Phase IV, no further attempts were made to operate the pilot system at the 7-day MCRT. It was decided that 5 days would be a more effective MCRT at which to operate. Phase V represents the first such operational period. However, more difficulty with sludge settleability ensued when the MCRT was reduced. During March 1976, techniques were developed for successfully stabilizing the system solids at design flow rates. Rather than using MCRT as an indication of stability, the sludge volume index (SVI) was used to determine when the solids were sufficiently stabilized to warrant step increases to the influent flow rate toward the design flow level.

By the end of March 1976, the system was operating very successfully and Phase VI, the first extended period of steady-state operation of the pilot plant, was begun. This became the most significant period of operation, since it showed conclusive evidence that the system could be maintained over long periods at design hydraulic loadings in a rectangular clarifier without violating discharge requirements for secondary effluent. It was during the latter part of this phase that the power reduction described earlier was made in the reactor's fourth stage. This period of good, steady operation was finally terminated on May 20, 1976, when repeated power outages, created by construction at the JWPCP, resulted in a pilot plant upset.

Continuing construction interruptions prevented a rapid return to steady operation. However, by June 21, 1976, the pilot plant was once again at steady-state conditions and a second sustained period of good operation (Phase VII) under design loading and conditions was begun. During this phase of operation, additional data were compiled relating both to organic and hydraulic parameters. Specifically, a series of radioactive tracer studies were begun during Phase VII which were designed to determine the movement of sludge solids through the final clarifier.

At the conclusion of the first series of clarifier tracer studies, the pilot plant was taken out of service and corrections were made to the baffle in the fourth stage of the reactor. This revision was outlined earlier in this report. After completing the baffle, operation was resumed in the longer of the two pilot clarifiers in order to accommodate further testing of sludge solids movement by the radioactive tracer method. Phase VIII summarizes the nearly 4 wk of operation in the long pilot clarifier, which represents the only change from operation during Phase VII.

Following the tracer studies, the flow was diverted back to the shorter clarifier and the diurnal flow pattern was again instituted. Some difficulties were encountered with the operation of the flow controller, but the pilot plant was stabilized in the diurnal flow pattern by October 28, 1976. Phase IX extended from October 28 to November 7, 1976, and was characterized by a 1900-m^3/day (0.5-mgd) average diurnally varied feed rate and a constant return sludge flow rate.

The clarifier operation during Phase IX was generally unsatisfactory. During the peak flow periods, the sludge blanket would rise to within 0.6 m (2-ft) of the surface, which resulted in an increase in effluent suspended solids. The peak flow in the diurnal cycle resulted in a clarifier overflow rate in excess of the design peak loading of 37-m^3/ m^2/day (900 gpd/ft^2), so on November 8, a 1900-m^3/day (0.5-mgd) peak flow diurnal flow pattern was introduced. Phase X extended from November 9 to November 24, 1976, and includes the data from the reduced diurnal flow pattern. During this period, the operation of the pilot plant improved, but the clarifier sludge blanket remained high during peak flow and the effluent suspended solids remained above the Federally-mandated 30 mg/l.

Further investigation indicated that the variation in the recycle ratio resulting from the constant return sludge flow and the diurnal influent flow was responsible for the poor clarifier performance. During low flow, the return ratio was high and the mixed liquor became more concentrated. When peak flow was reached, this concentrated mixed liquor was pushed into the clarifier and the clarifier loading was extremely high. This high solids loading was responsible for the high blanket and poor effluent.

To overcome this difficulty, it was necessary to operate the return sludge in a diurnal flow pattern. During Phase XI, December 10 to December 23, 1976, the same diurnal influent flow pattern employed in Phase X was used, but the return sludge flow was varied to maintain a constant recycle ratio. The return sludge had to be manually adjusted; therefore, because of manpower limitations, the pilot plant was operated at a constant flow of 1500 m^3/day (0.4-mgd) during the weekends. The frequent changes in operation modes caused minor upsets, but the pilot plant did produce satisfactory effluent quality.

SECTION 5

DISCUSSION OF RESULTS

EFFLUENT QUALITY

Activated sludge systems consist of two component units--the aerator/reactor and the final clarifier. The quality of the final effluent is related to the interaction of the component parts, and poor effluent may be caused by an inadequacy of only one part. The effluent quality of the air and oxygen systems is described in Tables 4 and 5.

Soluble COD and BOD

A primary indicator of the adequacy of the reactor in terms of oxygen transfer and treating the wastewater is the removal of soluble organics. In all phases, for both pilot plants, the soluble BOD_5 removals equalled or exceeded 95 percent. Phase average effluent soluble BOD_5 concentrations were 6 mg/l or less. These BOD measurements are low enough that differences between the two systems are not considered significant.

A small but definite difference between the systems is, however, apparent in the soluble COD data. The oxygen system produced effluent with consistently higher soluble COD. The data plotted in Figure 4 indicate that the principle cause of this is the lower aeration time maintained in the oxygen reactor. The oxygen data fit an eyed-in linear extrapolation to the air data reasonably well. The actual function should turn upward at the lower aeration times, reaching the influent concentration of 250+ mg/l at zero aeration time. Such a curve might be drawn to represent a better fit to the data in Figure 4.

When the soluble COD data are grouped according to aeration time and plotted against MCRT (Figure 5), it is apparent that, except at low values of less than 3-days, the MCRT has very little effect on soluble COD removal.

Suspended Solids

Secondary effluent solids concentrations depend on the effectiveness of the final clarifier. High effluent suspended solids, however, may be an indication of poor clarifier design, poor aerator design, or poor plant operation. During startup, both 1900-m^3/day (0.5-mgd) pilots plant experienced periods of high effluent suspended solids and turbidity, which were alleviated by reducing the power input to the final stages of the reactors.

TABLE 4. SUMMARY OF EFFLUENT QUALITY -- AIR SYSTEM

PARAMETERS	PHASE								
	I	II	III	IV	V	VI	VII	VIII	IX
Aeration Period (V/Q), hr	5 6	4.0	4 0	3 5	4 0	4.5	4 5	4 5	5 3
MCRT (Total System), days	6 8	5 4	3.3	5.6	2.8	4.3	4.3	4 5	5 9
Flow Pattern	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Steady
Suspended Solids									
Influent, mg/l	167	179	167	170	204	204	165	216	177
Effluent, mg/l	89	80	67	22	110	36	37	54	29
Removal, %	46.7	55 3	59.9	87 1	46 1	82.4	77.6	75 0	83 6
Total BOD_5									
Influent, mg/l 0	178	167	172	171	224	234	212	226	211
Effluent, mg/l 0	15	17	15	8	16	12	12	13	18
Removal, %	91 6	89 8	91 3	95 3	92 9	94 9	94 3	94 2	91 5
Soluble BOD_5.									
Influent, mg/l 0	118	102	98	101	126	132	129	109	119
Effluent, mg/l 0	2	3	3	4	5	4	2	2	2
Removal, %	98 3	97 1	96 9	96 0	96 0	97 0	98.4	98.2	98 3
Total COD:									
Influent, mg/l 0	458	447	453	460	513	556	483	515	517
Effluent, mg/l 0	118	152	130	77	191	91	92	111	84
Removal, %	74 2	66 0	71.3	83.3	62 8	83 6	81 0	78 4	83 8
Soluble COD									
Influent, mg/l 0	262	247	234	241	265	257	270	256	282
Effluent, mg/l 0	49	56	59	56	72	57	55	48	54
Removal, %	81.3	77 3	74 8	76 8	72 8	77 8	79.6	81 3	80 9
Grease (By Hexane Extraction)									
Influent, mg/l	51 2	40 6	36 5	37 8	-	-	-	-	-
Effluent, mg/l	8.2	6 1	4 8	1.0	-	-	-	-	-
Removal, %	84 0	85 0	86.8	97.4	-	-	-	-	-
Ammonia									
Influent, mg/l N	35.1	32 4	34.7	34 7	31 4	36.3	33.3	34 1	37 7
Effluent, mg/l N	14.2	20.3	27.8	31 6	27.8	32.1	27 5	32 1	30.7
Removal, %	59.5	37.3	19 9	8.9	11.5	11 6	17 4	6 1	18 6

TABLE 5. SUMMARY OF EFFLUENT QUALITY -- OXYGEN SYSTEM

PARAMETERS	PHASE I	II	III	IV	V	VI	VII	VIII	IX	X	XI
Aeration Period (V/Q), hr	2 5	3 1	3 4	2 8	2 5	2 5	2 6	2 5	2 5	3 1	3 0
MCRT (Total System), days	3 4	5 9	6 8	5 6	3 4	4 4	4 8	3 8	4 2	6 6	5 4
Flow Pattern	Diurnal	Steady	Steady	Steady	Steady	Steady	Steady	Steady	Diurnal	Diurnal	Diurnal
Suspended Solids											
Influent, mg/l	189	165	242	201	172	202	142	140	150	130	120
Effluent, mg/l	17	18	28	54	28	21	17	14	48	34	20
Removal, %	91 0	89 1	88 4	73 1	83 7	89 6	88 0	90 0	68 0	73 8	83 3
Total BOD_5											
Influent, mg/l 0	219	221	231	238	219	212	187	176	204	173	185
Effluent, mg/l 0	11	7	12	20	21	12	8	5	13	12	6
Removal, %	95 0	96 8	94 8	91 6	90 4	94 3	95 7	97 2	93 6	93 1	96 8
Soluble BOD_5											
Influent, mg/l 0	131	132	105	122	121	115	93	90	134	100	124
Effluent, mg/l 0	4	3	3	5	6	3	2	1	1	2	2
Removal, %	96 9	97 7	97 1	95 9	95 0	97 4	97 8	98 9	99 3	98 0	98 4
Total COD											
Influent, mg/l 0	467	523	554	561	486	536	438	400	415	431	446
Effluent, mg/l 0	81	87	94	122	100	88	82	71	116	97	83
Removal, %	82 7	83 4	83 0	78 3	79 4	83 6	81 3	82 2	72 0	77 5	81 4
Soluble COD											
Influent, mg/l 0	249	213	258	279	283	279	255	260	272	280	305
Effluent, mg/l 0	62	68	58	59	67	66	64	58	64	63	65
Removal, %	75 1	68 1	77 5	78 9	76 3	76 3	74 9	77 7	76 5	77 5	78 7
Grease (By Hexane Extraction)											
Influent, mg/l	42 6	38 4	47 1	55 8	41 6	62 4	63 8	45 8	46 0	39 2	40 6
Effluent, mg/l	1 0	0 9	3 0	4 4	2 5	1 7	1 6	1 3	6 2	2 8	2 3
Removal, %	97 7	97 7	93 6	92 1	94 0	97 3	97 5	97 2	86 5	92 9	94 3
Ammonia											
Influent, mg/l N	31 8	34 2	33 2	31 6	36 4	36 9	31 6	33 8	27 8	34 3	37 2
Effluent, mg/l N	26 3	31 4	30 5	31 3	31 0	31 5	29 5	28 9	28 0	28 7	33 8
Removal, %	17 3	8 2	8 1	0 9	14 8	14 6	6 6	14 5	-0 7	16 3	9 1

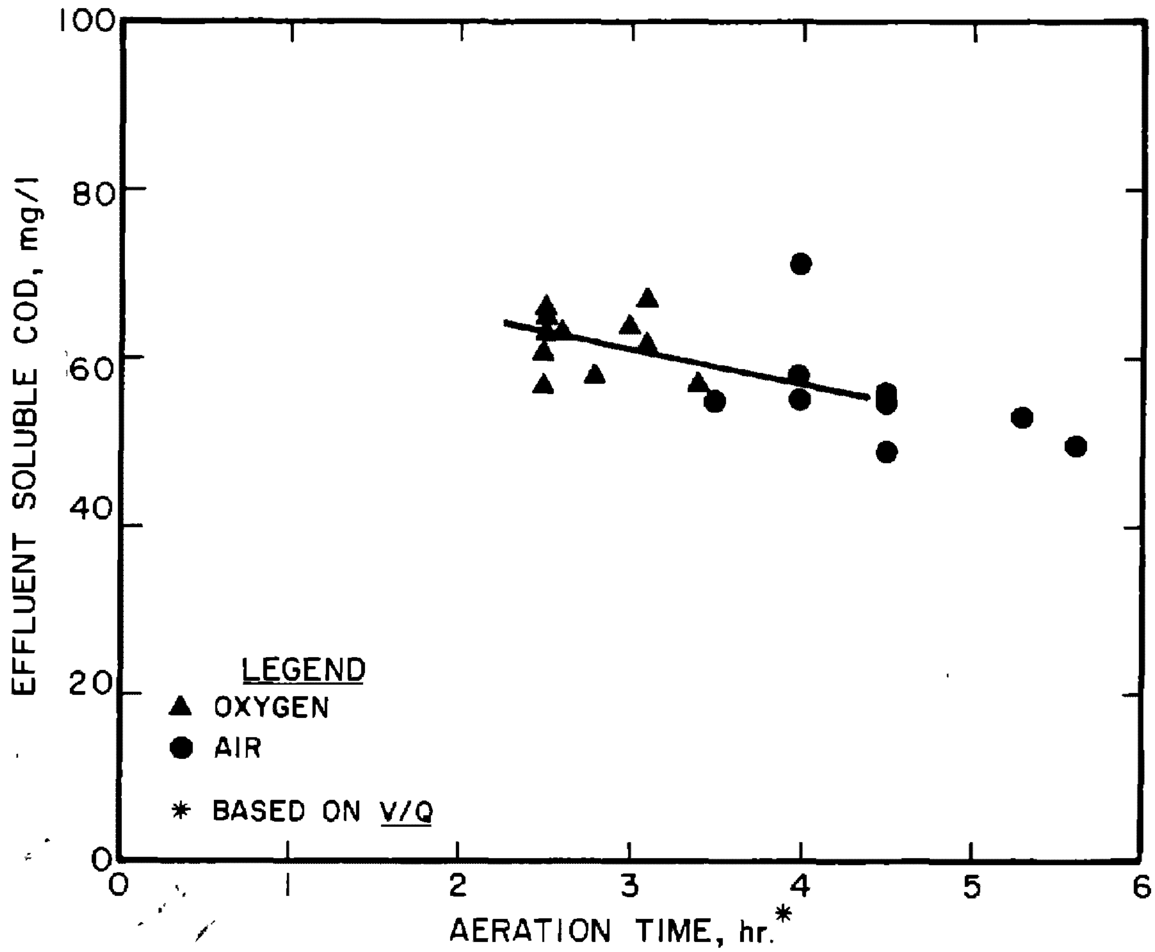

Figure 4. Soluble COD versus aeration time.

After startup, the oxygen system met the Federal discharge standard of not more than 30 mg/l for a 30-consecutive-day average in all phases except IV and IX. In both of these cases, the high effluent suspended solids can be traced to high clarifier solids loading. In Phase IV, the highest clarifier solids loading of the study, 168 kg/m^2/day (34 lb/ft^2/day), was experienced. During Phase IX, the average solids loading was lower, but during the peak of the diurnal flow pattern, the solids loading exceeded those in Phase IV. A major cause of the periodic high loadings in Phase IX rested in the return sludge operation. During Phase IX, the feed flow was varied in a diurnal flow pattern, but the return sludge flow was held constant. During low flows, the relatively high return sludge ratio would result in a concentrated mixed liquor in the reactor. When the influent flow was increased, the concentrated mixed liquor was forced into the clarifier at a high flow rate and corresponding high solids loading rate. Two steps were taken to correct this condition. The return sludge flow was varied in proportion to the influent flow to maintain a more nearly constant mixed liquor concentration, and the

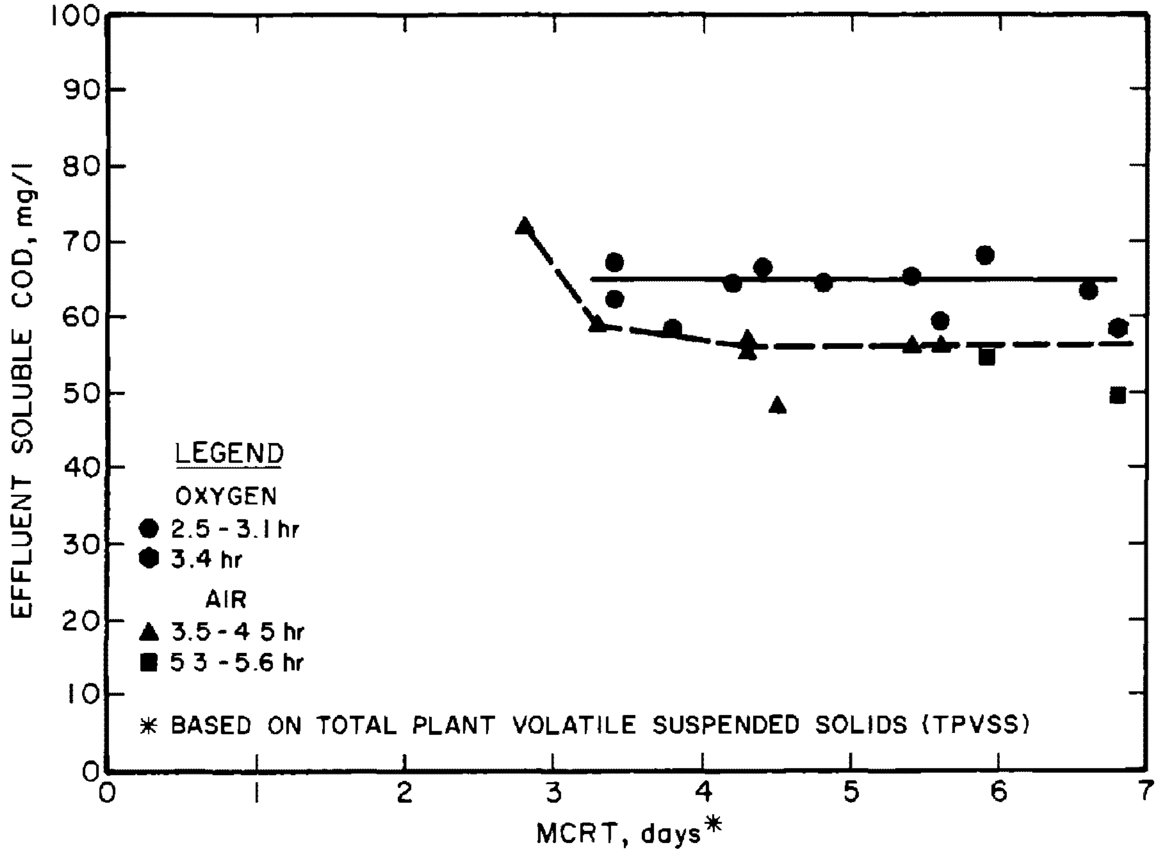

Figure 5. Effluent soluble COD versus MCRT.

flow to the unit was decreased from a 1900-m^3/day (0.5-mgd) average flow to a 1900-m^3/day (0.5-mgd) peak flow.

The deep tank submerged turbine system met the 30-mg/l effluent suspended solids standard only in Phases IV and IX. Phases VI and VII were characterized by generally low effluent suspended solids with a few unusually high days. Without those days, the 30-mg/l standard would have been met in those phases as well. The poorer performance of the air system is due in part to characteristics of air activated sludge and in part to the way the system was operated.

Three basic causes of high effluent suspended solids were observed during the DTST study. During startup and Phases III and V, the sludge did not flocculate and settle well. These conditions were attributed to excessive shear in the reactor during startup and the low MCRT's of 3.3 and 2.8-days maintained during Phases III and V, respectively.

Those low MCRT's were used to control the nitrification-denitrification that had occurred during Phases I and II. During the early part of the study, the DTST plant was operated in a manner conducive to partial nitrification. When the sludge was stored in the clarifier, nitrate and nitrite nitrogen were reduced to nitrogen gas. Bubbles formed which attached to sludge particles and resuspended them. Nitrifying bacteria grew more slowly than other activated sludge organisms, and the nitrification-denitrification conditions were eliminated by reducing the aeration time and/or the MCRT. In Phases III and V, however, the rising sludge was replaced by bulking sludge, and no improvement in effluent quality was achieved.

During Phases VI, VII, and VIII, the system was operated at the same aeration time and MCRT, but the recycle rate was varied from 30-percent to 40 percent and 50-percent. At the 30- and 40-percent recycle rates (Phases VI and VII, respectively) the pilot plant produced a generally good effluent, but at the 50-percent recycle rate (Phase VII), the clarifier was overloaded and the pilot plant produced poor effluent.

Although the oxygen pilot plant produced low suspended solids effluent more frequently than the air system, it is unfair to conclude from that information alone that oxygen activated sludge produces a lower suspended solids effluent. The oxygen system in these studies was operated much more conservatively than the air system. The oxygen system was operated within the known capability of such a system with an emphasis on refining certain design parameters, but the air system was operated to define the limitations of the deep tank turbine aeration system.

Both plants did demonstrate an ability to produce a good quality effluent. The air system, however, did prove to be more sensitive to operate. The main causes of this sensitivity is the tendency of the system to achieve partial nitrification, which resulted in rising sludge, and the measures that were necessary to control that condition.

Effluent Clarity

Clarity of an effluent is an aesthetic quality which is difficult to quantify. Since suspended solids greatly affect this quality, only periods with comparable effluent suspended solids concentrations can be used for comparisons. Those phases which averaged between 20 and 30 mg/l suspended solids were selected, and the data are presented in Table 6.

The turbidity in these effluent samples exhibited a correlation with suspended solids for each system, but the air system had slightly lower turbidities for given suspended solids concentrations. However, the Secchi disc transparencies, which were measured in the secondary clarifiers, indicate that the air system should have produced a much clearer effluent. Visibility in the final clarifiers was 20 to 40 percent greater in the air system. This confirms a general observation that whenever both systems were operating well, or rising sludge was present in the air system, the liquid fraction in the clarifier was much clearer in the air system than in the oxygen system. Similarly, the supernatant in the laboratory settling tests was visually much clearer for the air system than the oxygen system. No explanation for this is available at this time.

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TABLE 6. EFFLUENT CLARITY

SYSTEM	PHASE	SUSPENDED SOLIDS, mg/l	TURBIDITY, NTU	SECCHI DISC TRANSPARENCY, m (ft)
Air	IV	22	12	0.68 (2.2)
	IX	29	16	0.63 (2.1)
Oxygen	III	28	17	0.49 (1.6)
	V	28	21	0.44 (1.4)
	VI	21	14	0.55 (1.8)

Total COD and BOD_5

Since the secondary effluent suspended solids are primarily escaped biological floc, a direct correlation should exist between the effluent volatile suspended solids (VSS) and the effluent BOD_5 and COD. Cell material ($C_5H_7NO_2$) requires 1.42 times its mass in oxygen for complete oxidation.[1] If the effluent VSS are considered to be cell material, the nonfiltrable (soluble) COD and the nonfiltrable ultimate BOD will be 1.42 times the VSS. Figure 6 compares the nonfiltrable COD and ultimate BOD concentrations to the effluent VSS concentrations.

A least squares linear regression analysis was conducted on the oxygen COD data. The resulting line failed to pass through the origin, but the discrepancy was not statistically significant (40-percent confidence). The slope of the regression line was, therefore, adjusted to pass through the origin. A similar analysis was conducted on the air COD data, and the same line was established. For both systems:

$$\text{Total COD} - \text{Soluble COD} = 1.49\ \text{VSS}$$

The COD to VSS ratio of 1.49 is a reasonable experimental approximation of the theoretical 1.42 value.

1. Metcalf & Eddy, Inc., Wastewater Engineering, McGraw-Hill Book Company, New York, New York, 1972, p. 490.

TRATIONS

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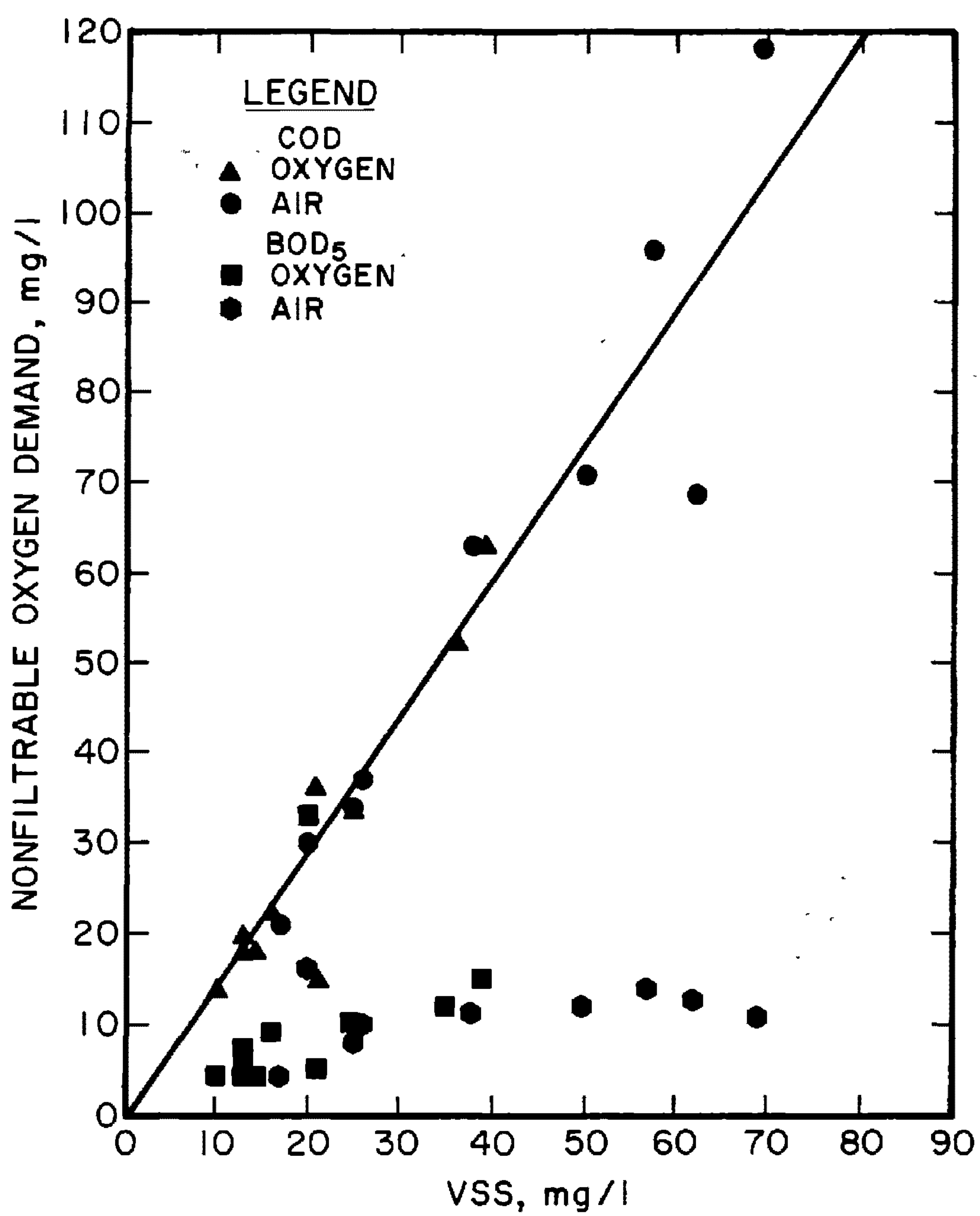

Figure 6. Nonfiltrable COD and BOD_5 versus VSS.

A visual inspection of Figure 6 indicates that the BOD_5 data do not represent a linear correlation with VSS. The highest-phase average total BOD_5 recorded for either system was 21 mg/l.

Hexane Extractables (Grease)

The phase-average effluent hexane extractable data from both 1900-m^3/day (0.5-mgd) pilot plants and the two smaller scale-pilot plants[2] are plotted against effluent suspended solids in Figure 7. A direct relationship between these two parameters is evidenced in Figure 7. The line drawn with an intercept at 0 and a slope of 0.086 is the linear regression of the data from the two large-scale systems. A linear regression of all data shown in Figure 7 yields a hexane extractable to suspended solids ratio of 0.067 and a soluble hexane extractable concentration background level of 0.9 mg/l.

While a theoretical relationship between grease and suspended solids has not been established to substantiate the experimental data, the importance of final clarification for grease removal has been emphasized. No difference in the grease removal efficiencies of the air and oxygen systems was found except the variation[6] caused by high effluent suspended solids.

Ammonia-Nitrogen

Four oxidation states of nitrogen are important in the operation of an activated sludge system. Nitrogen in wastewater is normally in the reduced state (-3). Reduced nitrogen is found free as ammonia or as a component of amino acids. In the presence of dissolved oxygen and specific bacteria, ammonia nitrogen may be oxidized to nitrite nitrogen (+3) and then to nitrate nitrogen (+5) through a process called nitrification. In a reducing environment, with the appropriate bacteria present, these oxidized forms may be reduced to elemental nitrogen (N_2 gas) by denitrification.

Ammonia nitrogen may be removed by nitrification or by conversion to cells. No indications of nitrificaton in the oxygen system were observed. Effluent nitrate and nitrite nitrogen were near zero, and a mass balance performed for the associated solids handling study indicated that all reduced nitrogen removal was due to cell synthesis.[3]

Low ammonia nitrogen removals are characteristic of most high purity oxygen systems since nitrification generally does not occur during the reaction process. There are usually two reasons given for this phenomenon: first,

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2. Stahl, J. F., Hayashi, S. T. Austin, S. R., Shamat, N., Summary Report - Operation of Small Scale Activated Sludge Pilot Plants at the Joint Water Pollution Control Plant, Los Angeles County Sanitation Districts, Whittier, California, April 1974.

3. Austin, S. R., Memorandum - Reduced Nitrogen Mass Transfer in the JWPCP Secondary Treatment System, Los Angeles County Sanitation Districts, Whittier, California, January 1978.

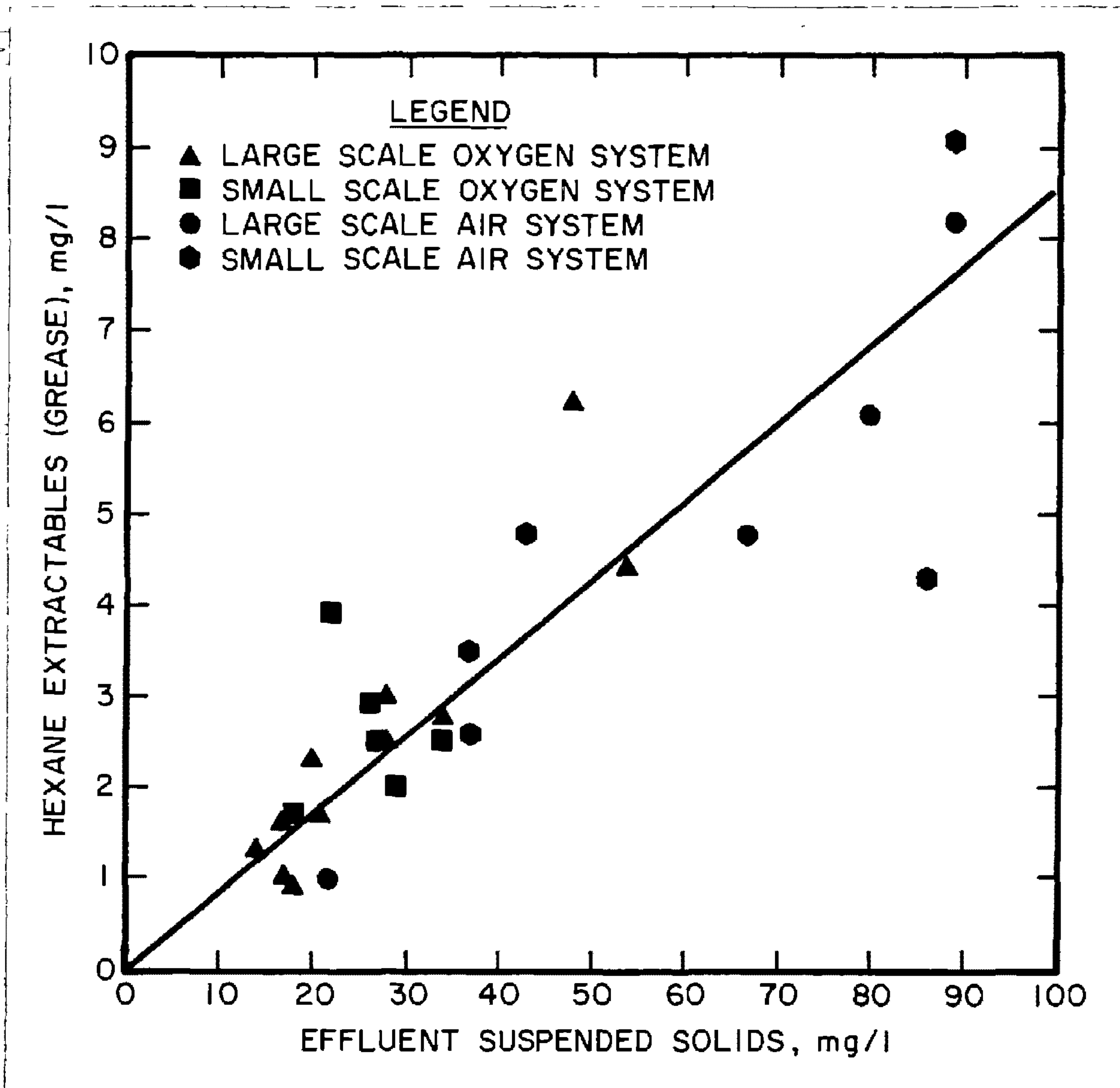

Figure 7. Hexane extractables versus effluent suspended solids.

pure oxygen systems are high-rate systems, which usually means that the process is operated at a low MCRT, thus reducing the possibility that nitrifying organisms will establish themselves within the biomass; second, most pure oxygen systems are sealed reactors to maximize oxygen utilization. As a result, there is a reduction in the pH of the mixed liquor through the dissolution of carbon dioxide which further inhibits the growth of nitrifying bacteria.

Ammonia nitrogen removals up to 60 percent were observed in the air system. In fact, controlling nitrification was a major consideration in the operation of the air pilot plant. Nitrification followed by denitrification caused rising sludge and high effluent suspended solids.

In order to control nitrification, the aeration time and MCRT were reduced. The ammonia nitrogen removals after Phase II may be attributed to cell synthesis.

Trace Metals, Cyanide, and Phenols

Certain trace constituents were monitored during the activated sludge studies at the JWPCP. The data from the 1900-m^3/day (0.5-mgd) pilot plants confirmed the data from the small-scale systems, so a reduced sampling schedule was employed on the larger systems. The data from all four systems are presented in Tables 7 and 8. The discharge limitations imposed by the California Regional Water Quality Control Board (RWQCB) on the JWPCP are also included.

Chromium, nickel, and zinc are the three trace constituents that are in violation of the RWQCB standards and will require source control in the Joint Outfall System. Removals of these metals were similar in the four activated sludge systems with 67 to 74 percent of the chromium, 23 to 50 percent of the nickel, and 55 to 68 percent of the zinc being removed.

The influent arsenic concentrations were near the detection limit, so the removal data are of minimal value. Removals of the other metals ranged from 40 to 83 percent, with neither the air nor the oxygen systems having a clear advantage.

Cyanide and phenols are organic complexes that are subject to oxidation. Cyanide removals ranged from 64 to 86 percent with the oxygen system obtaining the higher removals. Removal of phenols was 98 percent or higher, with the air system producing effluents at or below the detection limit.

SLUDGE PRODUCTION

One of the most important claims made on behalf of pure oxygen is that the net growth of solids in these systems will be less than a similar air system when operated at the same MCRT. Since a large portion of the cost of wastewater treatment is usually associated with solids processing and sludge handling, this claim would represent a significant savings in both capital and operating costs. The claim is based on a comparison between the two

TABLE 7. TRACE CONSTITUENT REMOVAL BY MEANS OF AIR-ACTIVATED SLUDGE

Constituent	RWQCB Standard		1900-m^3/day (0 5-mgd) Pilot Plant			1 6-1/sec (25-gpm) Pilot Plant			
	Average, mg/l	10% of Time, mg/l	Influent, mg/l	Effluent, mg/l	Removal, %	Influent, mg/l	Effluent, mg/l	Removal, %	% Samples Over 10% Standard
Arsenic	01	02	01	01	0	02	01	50	0
Cadmium	02	03	017	008	53	020	008	60	0
Total Chromium	005	--	28	08	71	47	14	70	100
Copper	20	30	22	06	73	33	11	67	2
Lead	10	20	15	06	60	15	06	60	0
Mercury	001	002	--	--	--	0007	0003	57	0
Nickel	10	.20	.26	18	31	30	23	23	63
Silver	02	.04	010	006	40	012	005	58	0
Zinc	.30	50	1.36	0 58	57	1 43	46	68	35
Cyanide	10	.20	0.14	0 05	64	38	08	79	3
Phenols	50	1.00	2 88	0 01	99+	1 41	01	99+	0

TABLE 8. TRACE CONSTITUENT REMOVAL BY MEANS OF OXYGEN-ACTIVATED SLUDGE

Constituent	RWQCB Standard		1900-m^3/day (0 5-mgd) Pilot Plant			0 6-1/sec (10-gpm) UNOX Mobile Pilot Plant			
	Average, mg/l	10% of Time, mg/l	Influent, mg/l	Effluent, mg/l	Removal, %	Influent, mg/l	Effluent, mg/l	Removal, %	% Samples Over 10% Standard
Arsenic	01	02	--	--	--	02	01	50	0
Cadmium	02	03	* 024	* 007	*71	024	004	83	0
Total Chromium	005	--	27	07	74	46	.15	67	100
Copper	20	30	22	06	73	35	06	83	0
Lead	10	20	* 14	* 03	*79	15	08	47	0
Mercury	001	002	--	--	--	0007	0002	72	0
Nickel	.10	20	23	15	35	.30	23	23	80
Silver	02	04	* 015	* 004	*73	.013	.003	77	0
Zinc	30	50	1 01	40	60	1 27	57	55	55
Cyanide	.10	20	--	--	--	35	.05	86	0
Phenols	50	1 00	--	--	--	1 62	03	98	0

*Results of only one analysis

systems that shows the net sludge production (VSS produced/COD_R) of air systems to be greater for any given organic loading rate (COD_R/MLVSS) than a similarly operated oxygen system.

From an analysis of the data collected both from the small- and large-scale units the Districts have concluded there is little difference between the air and oxygen systems in terms of sludge production. When an analysis of the system is made based on the mass of microorganisms contained within the biological reactor (which is the method used by proponents of pure oxygen), the data does indeed indicate that the oxygen system produces less sludge. It is the belief of the authors, however, that the mass of solids within the entire biological system must be considered in order to obtain a true indication of the level of sludge production. This means that the solids that are present in the final clarifiers must be included when the total system solids are calculated. When the data is re-examined in this way, the oxygen system will no longer demonstrate an advantage over air systems in terms of sludge production. This reversal is due to the fact that a greater portion of the total system solids will be contained within the clarifiers of an oxygen system than is typically encountered in air-activated sludge systems. As was outlined earlier, improved sludge settling and oxygen transfer capability allows the oxygen system to be operated as a high-rate system. As a result, as much as 50 percent of the total system solids will be carried in the final clarifiers. If the air and oxygen systems are compared based on reactor solids only, then a significant portion of the oxygen solids will be eliminated form the analysis, thus falsely indicating a higher organic loading rate than that imposed on the air system.

A sludge growth kinetics analysis based on total system solids is presented in Figure 8. Linear regression lines (developed by treating the MCRT as the independent variable) are shown for the air and oxygen data along with the 90-percent confidence limits for the location of the oxygen line. It is not possible to reject, with 90-percent confidence, any line falling within these limits as the true line from which the oxygen data were generated. Since the air system regression line falls within these confidence limits, the oxygen growth kinetics are not distinct from the air kinetics at the 90-percent confidence level.

The observed net sludge production data (which includes the VSS in the waste sludge plus the effluent) are plotted as points in Figures 9 and 10. Identical data are presented in both figures. The graphic display of the data points shows that it is difficult to determine which system has a higher net sludge production.

The curves superimposed on the data in Figures 9 and 10 were developed from the growth kinetics shown on Figure 8. The two linear regression lines for the air and oxygen system shown on Figure 8 (developed using the MCRT as the independent variable) are:

$$1/\theta_c = 0.58\ (F/M) - 0.21 \qquad \text{Air System}$$

$$1/\theta_c = 0.81\ (F/M) - 0.32 \qquad \text{Oxygen System}$$

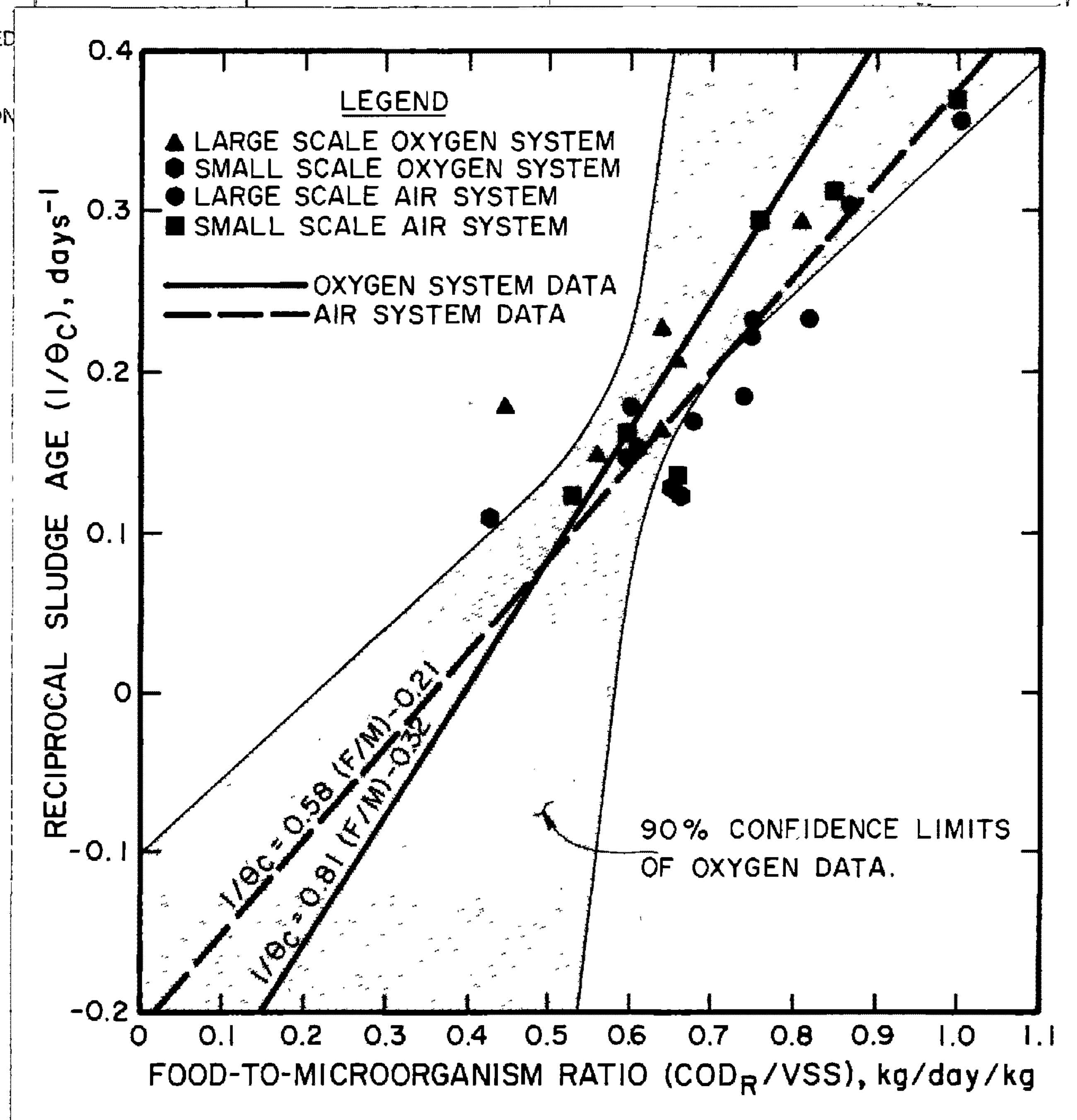

Figure 8. Sludge growth kinetics.

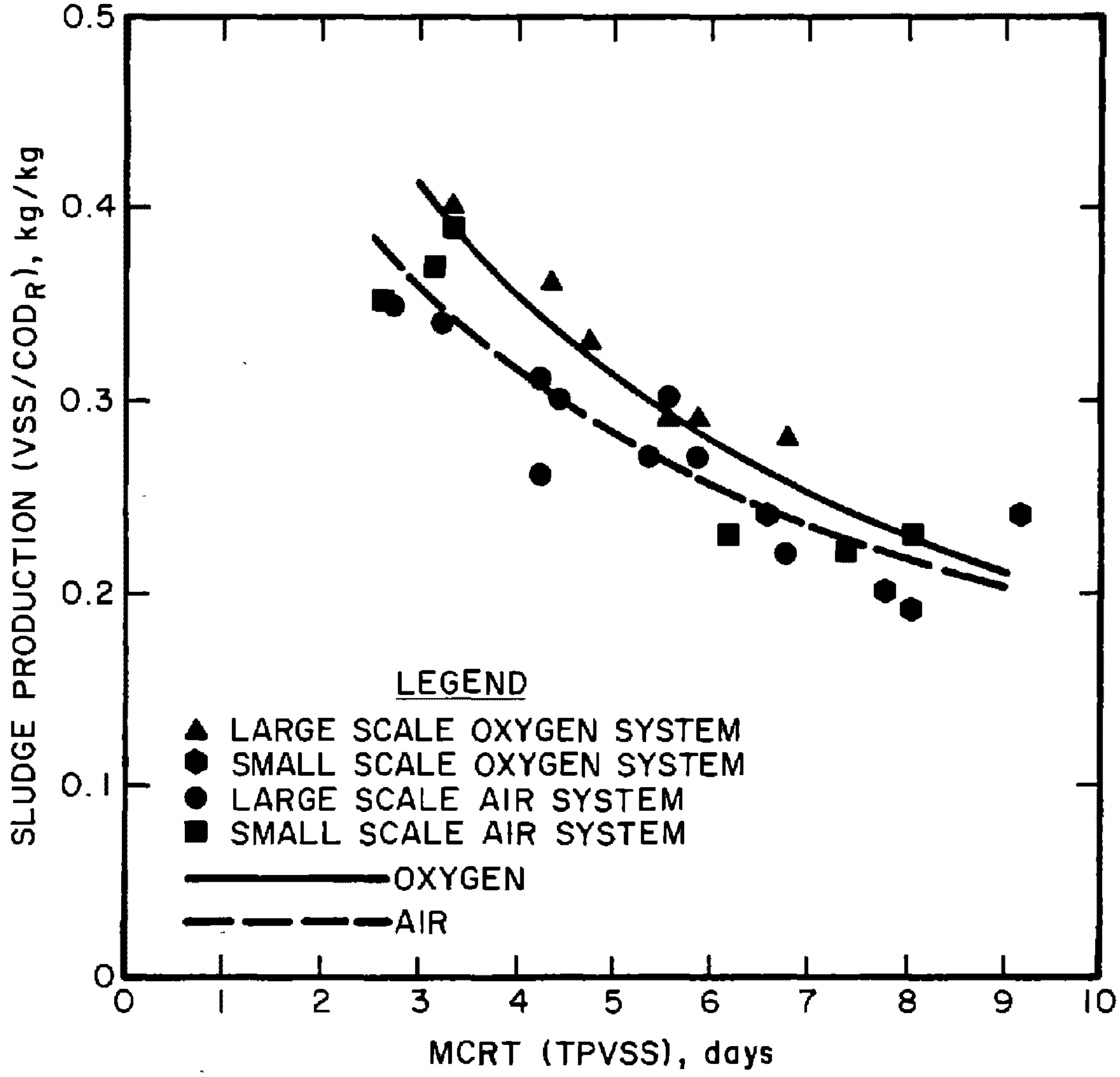

Figure 9. Analysis of net sludge production using MCRT as the independent variable.

These equations are in the form of

$$1/\theta_c = Y\ (F/M) - k_d$$

where:

θ_c = MCRT

F/M = food-to-microorganism ratio

Y = growth yield coefficient

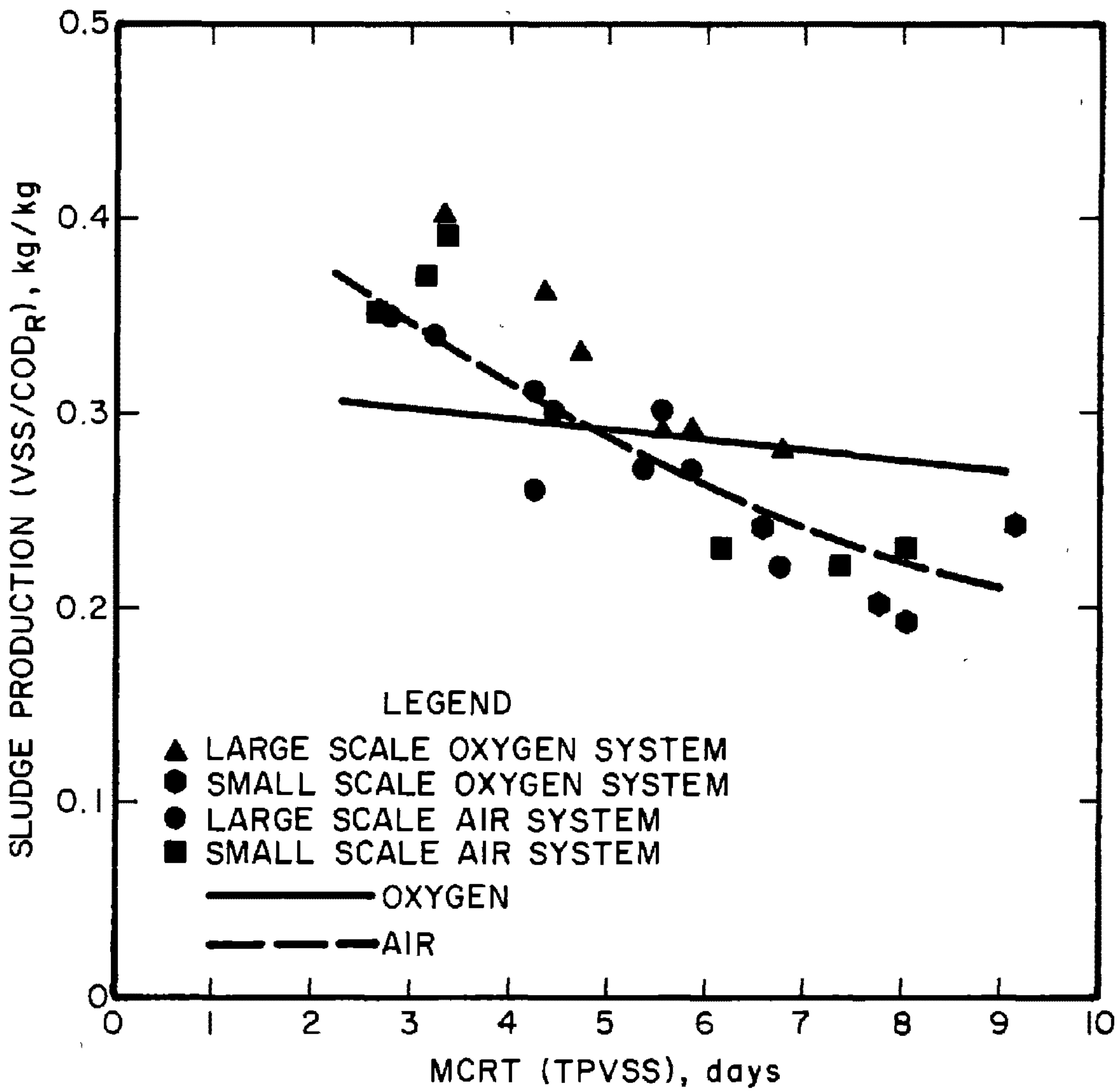

Figure 10. Analysis of net sludge production using the food-to-microorganism ratio as the independent variable.

k_d = microorganism decay coefficient

The net sludge production (VSS/COD_R in kg/kg) is defined as follows:

$$\text{Net sludge Production} = Y/(1 + k_d\theta_c)$$

The air and oxygen curves shown in Figure 9 were derived using the above formula for net sludge production with Y and k_d being supplied from the linear regression analysis in Figure 8. Though the linear regression lines in Figure 8 were shown to be statistically insignificant at the 90-percent confidence limits, it is interesting to note that the curves in Figure 9 indicate that the oxygen system has a higher net sludge production.

A linear regression analysis was also conducted on the data in Figure 8 using F/M as the independent variable, rather than the MCRT. The linear regression analysis assumes that the independent variable is exact and adjusts the line to best fit the data. Therefore, using F/M as the independent variable rather than the MCRT produces slightly different lines than those shown on Figure 8. The linear regression lines produced using F/M as the independent variable are:

$$1/\theta_c = 0.50\ (F/M) - 0.15 \qquad \text{Air System}$$

$$1/\theta_c = 0.32\ (F/M) - 0.02 \qquad \text{Oxygen system}$$

The air and oxygen curves presented in Figure 10 were derived using the previously given formula for net sludge production with Y and k_d being supplied from the linear regression lines given above. This analysis shows that at MCRT's of above 5 days, the oxygen system again has a higher net sludge production than the air system.

Because of reservations regarding mass balances, the data form the last four phases of the oxygen system operation have not been used. On both Figures 9 and 10, those data would have tended to move the oxygen system sludge production curve upward at the lower MCRT's.

SLUDGE SETTLEABILITY

Two parameters are commonly used to indicate sludge settleability. The sludge volume index (SVI) is the inverse of the settled sludge concentration expressed in ml/g, and the initial settling rate (ISR) is the maximum rate at which the sludge interface drops during the test.

The 30-min SVI data were presented previously in Tables 2 and 3. The phase-average oxygen system SVI varied from 65 to 153 ml/g, with an average of 99 ml/g, and the air system produced SVI's of 146 to 252 ml/g, with an average of 167 ml/g.

The ISR data resulted from one series of tests which was conducted during a period when the performance of both pilot plants was charcterized as "good." In this series of tests, the oxygen sludge settled about three times as fast as the air sludge (Figure 11). These are the results of only one test, but they are in qualitative agreement with the general experience at the JWPCP.

The oxygen sludge definitely settles better and gravity thickens better than the air sludge. However, it is not possible at this time to determine the extent to which this is an innate property of oxygen-activated sludge or a function of the reactor design.

One factor which affected the sludge settleability in both of these systems was the power input. During the startup of each pilot plant it was necessary to reduce the mixer power in order to produce an acceptable effluent. Excessive power input shears the floc, which can cause poor settleability of the sludge and a turbid effluent.

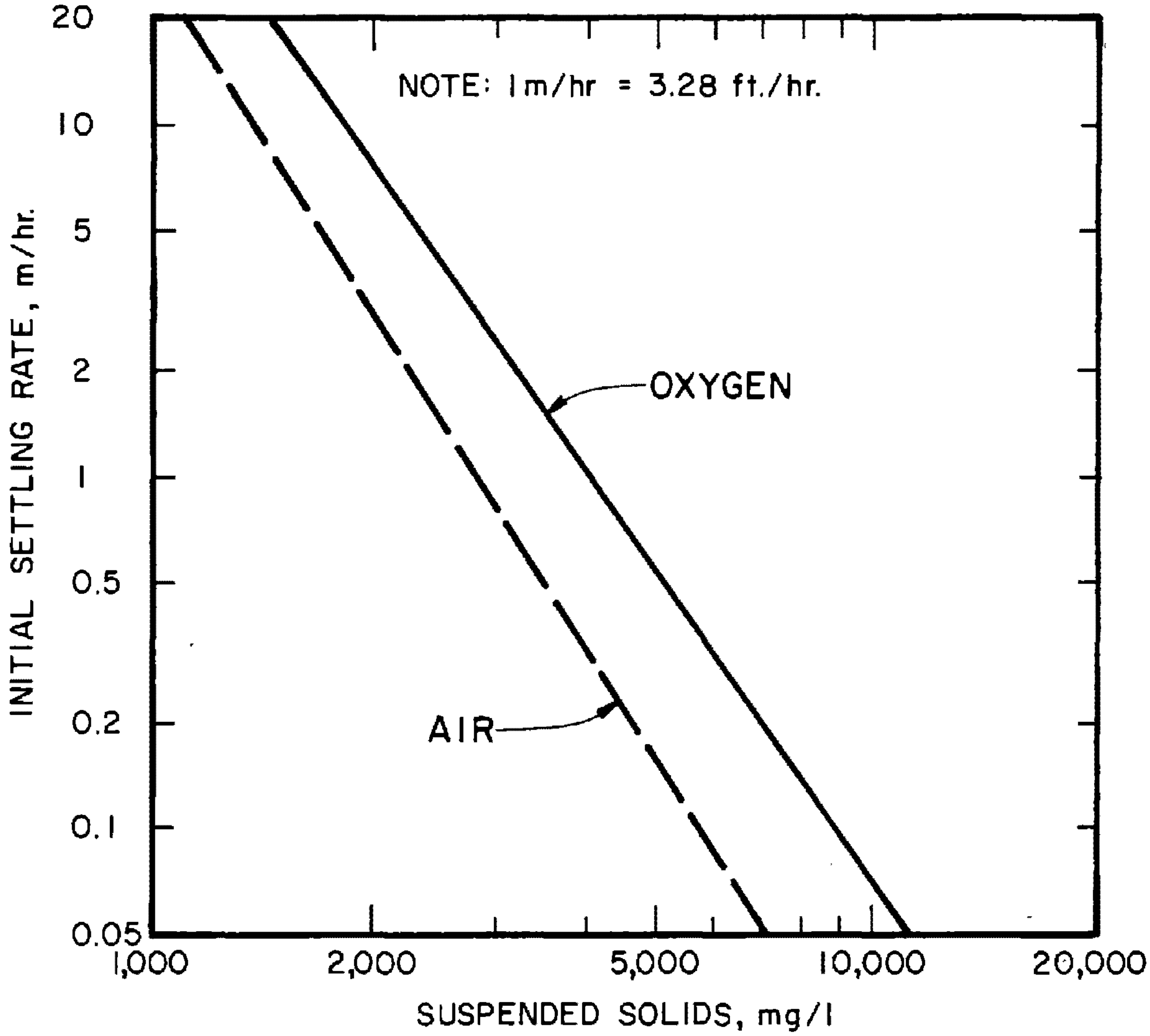

Figure 11. Initial settling rates.

POWER CONSUMPTION

In the present economic climate, one of the most important factors involved in the comparison of air- and oxygen-activated sludge processes concerns energy consumption. Since power intensity problems in both pilot plants required the aeration equipment to be operated at speeds lower than design, a comparison based on the pilot plant data is inappropriate. Additionally, the effects of scale would be difficult to predict, so estimates based on typical aerator efficiencies will produce more applicable results.

The standard oxygen transfer rates (SOTR's) presented in Table 9 are representative of present mechanical aeration technology, although specific equipment may differ from those values. In the case of the submerged turbine, a 50:50 power split between the mixer and compressor was assumed. As indicated in

Table 9, different mechanical efficiencies were assigned to the mixer and compressor to obtain wire power consumptions.

The oxygen transfer equation is:

$$\frac{dC}{dT} = \alpha K_L a \; (\beta C^{*} - C) \qquad (1)$$

where:

$\frac{dC}{dt}$ = oxygen transfer rate, mg/l/hr

$K_L a$ = volumetric mass transfer coefficient, hr^{-1}

C^{*} = equilibrium dissolved oxygen concentration at zero uptake, mg/l

C = system dissolved oxygen concentration, mg/l

α, β = variables to correlate clean water results to mixed liquor conditions.

By adding a power intensity term (V/P), it is possible to obtain an equation in which the left side has the same units as the SOTR.

$$SOTR = \frac{dC}{dt}\left(\frac{V}{P}\right) = \alpha \left[\frac{K_L a \; V}{P}\right] (\beta C^{*} - C)$$

where:

V = tank volume ($10^3 m^3$)

P = power (kW)

It is now possible to apply the standard conditions in Table 9 to obtain the constant $K_L aV/P$ and then determine the oxygen transfer rate under field condition.

The air system was a completely mixed reactor with dissolved oxygen (DO) maintained at 1 mg/l. Using an equivalent depth (the depth associated with a saturation DO of C^{*}) at 0.4 of the air introduction depth, and the conditions listed in Table 9, the calculations are straightforward.

Since the oxygen system is muilti-staged, the model is slightly more complicated. Based on the conditions observed in pilot studies (DO, gas purity, and oxygen uptake rate, see Table 10) and communications with manufacturers, a model was developed which allowed the calculation of required $K_L a$'s in each stage. Since $K_L a$ is proportional to power, the data allow the power fraction in each stage to be calculated. The transfer efficiency at field conditions in each stage was calculated, and an average based on the power distribution provides the overall efficiency. The power required to extract the pure

TABLE 9. ASSUMED OXYGEN TRANSFER RATES

System	Standard Oxygen Transfer Rate[a] (Delivered Power) kg/kWh	(lb/hp-hr)	Power Consumption (Delivered) kWh/kg (hp-hr/lb)	Power Transfer Efficiency	Power Consumption (Wire) kWh/kg (hp-hr/lb)	Oxygen Transfer Rate[a] (Wire Power) kg/kWh (lb/hp-hr)
Surface Aerator	2 13	(3 50)	0 469 (0 285)	0 839[b]	0 559 (0 340)	1 79 (2 94)
Submerged Turbine Total System	1 70	(2 80)	0 588 (0 348)		0 830 (0 505)	1 20 (1 98)
Mixer			0 294 (0 179)	0 839[b]	0 350 (0 213)	
Compressor			0 294 (0 179)	0 612[c]	0 480 (0 292)	

[a] Standard Conditions Gas Purity = 21% O_2, Water Temperature = 20°C, Dissolved Oxygen = 0 mg/l, α = 1 00, β = 1 00

[b] Efficiencies Gear Box = 0 96, Coupling = 0 95, Motor = 0 92

[c] Efficiencies Blower = 0 70, Coupling = 0 95, Motor = 0 92

TABLE 10. OXYGEN SYSTEM OPERATING CHARACTERISTICS

Stage	Gas Purity % O_2	Dissolved Oxygen, mg/l	Power Fraction
1	80	9 1/3	0.46
2	70	7	0.27
3	65	5	0.15
4	50	2	0.12

oxygen from the atmosphere must be added to the aerator power in order to provide a fair comparison.

The results of these calculations are presented in Table 11. The oxygen systems use substantially less energy in this analysis. The surface aerator oxygen system, in fact, is estimated to require only 52-percent of the energy used by the air system, and the submerged turbine oxygen system is projected to need 62-percent of the energy used by the air system. Because of land constraints at the JWPCP, depths greater than 5-m (15-ft) were required for the air system, so surface aeration was not evaluated for the air system

DEPENDABILITY AND MAINTENANCE

In the JWPCP studies, the oxygen-activated sludge process has proven to be very stable and has generally recovered from upsets very quickly. The major operational problems have been associated with the appurtenant equipment, which is much more complex than is encountered in most air systems. Because of the potential for explosions in the enriched atmosphere, oxygen-activated sludge systems must be equipped with an explosive vapor detector. This equipment has proven subject to frequent failures, which have automatically shut down the total aeration system.

One maintenance item that has not been quantified, and had not been expected, concerns life of the clarifier flight chains. The oxygen effluent has proven to be much more aggressive to the cast links than the air effluent. This is probably a result of the higher dissolved oxygen content and the lower pH of the oxygen effluent.

TABLE 11. POWER CONSUMPTION

System	Aerator Type	Water Depth, m (ft)	Power Consumption (Wire Power), kWh/kg O_2 transferred, (hp-hr/lb O_2 transferred)		
			Aeration Equipment[a]	Oxygen Generation[b]	Total
Air	Submerged	7.6 (25)	1.28 (0.78)		1.28 (0.78)
Oxygen	Submerged	4.6 (15)	0.44 (0.27)	0.35 (0.21)	0.79 (0.48)
Oxygen	Surface	4.6 (15)	0.31 (0.19)	0.35 (0.21)	0.66 (0.40)

[a] Turbine plus compressor: Water Temperature = 23 C, α = 0.80, β = 0.95.

[b] Based on JWPCP design, 90% oxygen utilization.

As a biological process, the air system seemed to be more sensitive than the oxygen system. Due to the air system's tendency to nitrify and the associated rising sludge, it was necessary to operate the air system at low aeration times and low MCRT's. Operating in this marginal region has contributed to the sensitivity of the air system.

Mechanically, the air system was much simpler and less subject to malfunctions than the oxygen system.