APPLICATION OF OXYGEN TO TREAT WASTE FROM MILITARY FIELD INSTALLATIONS

AN EVALUATION OF AN ACTIVATED SLUDGE PROCESS EMPLOYING DOWNFLOW BUBBLE CONTACT AERATION

Final Report
to the
United States Army
Medical Research and Development Command
Preventive Medicine Research Branch

Washington, D.C.

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July, 1974

EHE-74-01

CRWR - 110

ACKNOWLEDGEMENTS

The City of Austin, and in particular the Department of Waster and Waste Water Treatment, and the Chief Engineer, Mr. M.W. Brenerman, should be thanked for their kind permission to conduct research at Govalle Treatment Plant. The authors express their thanks to Mr. R. A. Pfaffman, supervisor of Govalle Treatment Plant for his assistance and interest.

In addition our appreciation is extended to Mr. Mauricio Guenero, whose responsibility was assistance in operation and maintenance of the pilot system, and Mr. Alan Barnett, whose responsibility was for analytical support.

Dr. R. E. Speece, Dr. J.O. Ledbetter, and Dr. G.A. Rohlich provided valuable advice, comments, and review.

This project was supported by The United States Army, Medical Research and Development Command, Preventive Medical Research Branch under Contract DADA 17-72-C-2154.

ABSTRACT

A pure oxygen system incorporating Downflow Bubble Contact Aeration as the transfer unit was operated utilizing settled municipal wastewater as feed. Oxygen transfer efficiencies in excess of 95 percent were obtained.

The kinetic coefficients were determined and compared favorably with previously reported pure oxygen values. The yield was 0.47 lb volatile suspended solids per lb COD removed and the decay coefficient was 0.02 per day. The maximum specific growth rate was 3.7 lb VSS produced per lb VSS and the half-velocity constant was 281 mg/l.

The BOD removal was 94.5 percent at substrate removal rates of up to 0.73 lb BOD per lb MLVSS per day. The system removals are as good as any previously reported pure oxygen values.

Solids separation appeared to be excellent with an SVI of 84 ml/g. This compared quite well with other reported sludge data for high purity oxygen activated sludge systems. Auxiliary mixing appeared to be necessary in order to obtain good sludge separation.

TABLE OF CONTENTS

page

Chapter

TABLE OF CONTENTS CONTINUED

TABLE OF CONTENTS CONTINUED

LIST OF TABLES

LIST OF FIGURES

LIST OF FIGURES CONTINUED

I. INTRODUCTION

The use of high purity oxygen in the activated sludge process has increased in recent years. Oxygen systems are technically feasible in many cases but often not economical. Most operational problems have been rectified, but economical oxygen transfer has not been completely attained.

Oxygen systems have been troubled by economics from their inception by Okun, who had hoped that savings in capital expenses realized from smaller plants could offset the high cost of oxygenation (6). However, the high cost of pure oxygen in the early 1950's prevented oxygen systems from competing with conventional air units.

In the late 1960's interest in the pure oxygen process was revived largely as a result of cheaper oxygen production methodology (10). After several years of operation at Batavia, New York, Union Carbide concluded that substrate removal rates were higher, sludge yields were lower, and endogenous respiration rates were higher in oxygenated systems than in aerated tanks. Projected economics showed pure oxygen to be less expensive than air; however, oxygen transfer was a major cost. Aeration tanks were covered with tight fitting covers, and large mixers and compressors were employed to improve oxygen transfer.

Later modifications of the UNOX Process such as those at Newtown Creek, Blue Plains, and Simplox all maintained the costly covered tank for oxygen transfer (12, 14). Of course, each system utilized a slightly different approach to oxygen injection, from spargers to diffusers.

Oxygen transfer efficiencies of up to 95 per cent have been reported, but the corresponding transfer costs (covered tanks and equipment) have offset the considerable savings from smaller tank sizes that Okun had foreseen. Speece et al. (15) developed the downflow bubble contact aeration system, which is capable of oxygen absorption efficiencies of well over 90 per cent in a compact unit. The system may be pressurized and dissolved oxygen concentrations of over 100 mg/l are possible and economical. Therefore, this device was used in a pure oxygen system as the transfer unit. Accordingly, it has become necessary to investigate the operational characteristics and to determine the treatment efficiencies and kinetics of a pure oxygen system using Downflow Bubble Contact Aeration.

II. SCOPE OF INVESTIGATION

The principal objective of this investigation was to evaluate the performance of the Downflow Bubble Contact Aeration pure oxygen system (DBCA System). Specifically, the treatment capabilities of the DBCA system were determined and the results compared with data previously reported for oxygen systems. Also, the operational characteristics and problem areas inherent in present DBCA system design were identified.

III. LITERATURE REVIEW

ROLE OF OXYGEN IN BIOLOGICAL TREATMENT

Molecular oxygen is required in aerobic biological processes as the final electron acceptor during the aerobic oxidation of substrates by microorganisms. The molecular oxygen acts as the ultimate sink for the electrons removed from reduced organic materials in wastewaters. The energy released during oxidation is utilized in two basic processes: cell synthesis and endogenous respiration (1). Thus, one of the most important requirements of a well-operated aerobic biological treatment process is the continuous and consistent maintenance of an adequate dissolved oxygen supply. A process objective, therefore, is to maintain sufficient dissolved oxygen levels to satisfy the bacterial uptake requirements.

Gaden (2) found that oxygen uptake rates were drastically reduced below a critical dissolved oxygen level of 0.3 mg/l. Hence, in order to operate above the critical level throughout the aeration basin, practical operating experience has indicated that a minimum of one to two mg/l of dissolved oxygen must be maintained (3).

In order to provide for the maintenance of the desired dissolved oxygen levels in the system, the typical approach has been to develop more efficient transfer equipment. However, the driving force of oxygen-solution interchange is proportional to the oxygen deficit. Even with the most efficient equipment, the practical maximum transfer of oxygen from air into solution at atmospheric pressure is no more than seven mg/l (4).

Another approach has been to increase the driving force for oxygen transfer by providing a high purity oxygen concentration in the gas phase. The driving force for oxygen transfer is raised to about 40 mg/l when high purity oxygen is substituted for air because the solubility of oxygen in water at normal temperatures and pressures is approximately 43 mg/l (5).

SIGNIFICANT RESEARCH WITH HIGH PURITY OXYGEN

Bio-precipitation

Bio-precipitation, a biological treatment system which utilized high purity oxygen was investigated by D. A. Okun in 1940 (6). The bio-precipitation process shown in Figure 1 consisted of a combination reactor-clarifier and an external oxygen absorber. Influent and recycled effluent flowed down through the absorber and was pumped to the bottom of the reactor clarifier. Influent was distributed across the unit and then passed up through the sludge blanket in the reactor section. Stagnant areas were prevented by slow moving mixers. Treated effluent separated from the sludge in the clarifier section and was discharged or recycled back

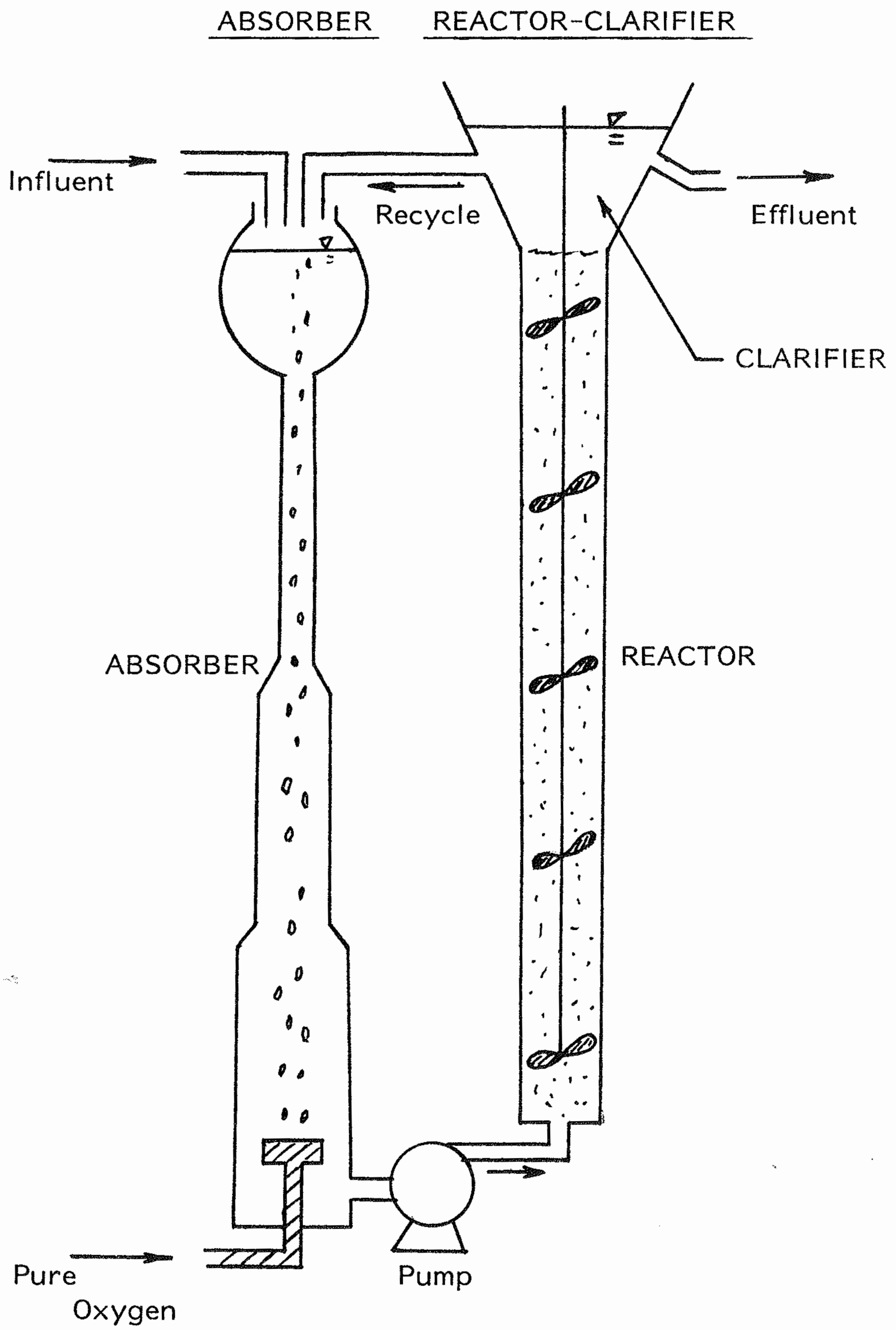

FIGURE 1. BIO-PRECIPITATION PROCESS

to the absorber.

The absorber and reactor-clarifier were open to the atmosphere allowing oxygen to escape and nitrogen to enter the absorber, thereby decreasing its efficiency. To provide enough oxygen in the reactor-clarifier, a recycle ratio of clarified effluent to influent was maintained at five to one. The dissolved oxygen concentration in the influent to the reactor was about 30 mg/l while the effluent concentration was about seven mg/l. Two units were constructed for comparative studies under air and oxygen conditions. Unfortunately, a true comparison was not possible since up to 17 times as much recycled clarified effluent per unit of influent was required under air operation. The settling properties of the sludge under such conditions were affected.

The laboratory results reported by Okun did not include a comparison of sludge yields for the parallel oxygen and air units. However, sufficient data were available to reevaluate the substrate removal rate. The relationships between substrate removal rate in pounds of BOD removed per day per pound of mixed liquor suspended solids and effluent BOD for the air and oxygen systems are illustrated in Figure 2. The usual result of such a plot is a straight line at low substrate concentrations (8). The slope of the line indicates the activity of the biomass. The data from both the air and oxygen systems falls on the same line; therefore, no significant difference

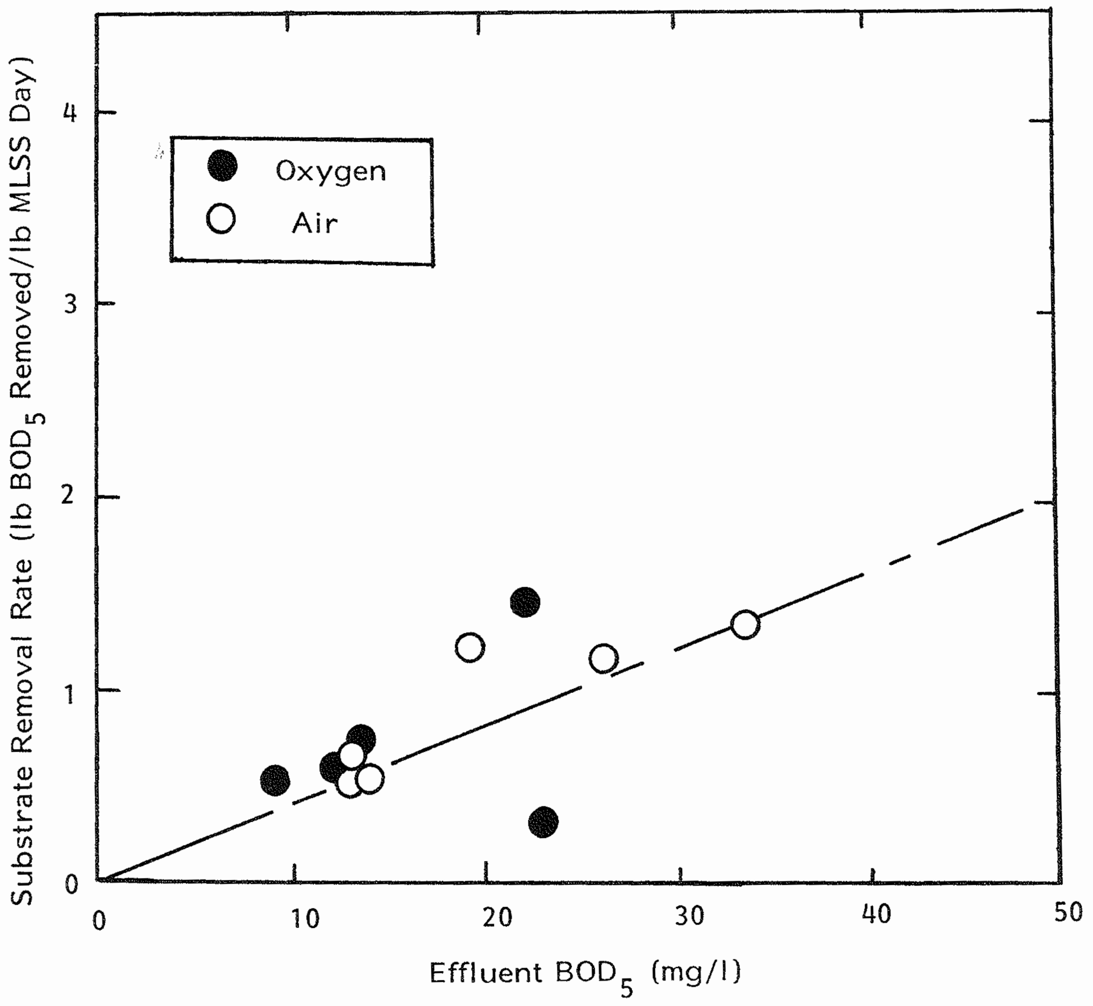

FIGURE 2. RELATIONSHIP BETWEEN SUBSTRATE REMOVAL AND EFFLUENT BOD_5, BIO-PRECIPITATION

existed between the metabolic activity of organisms in the air and high purity oxygen. These relationships agree with the conclusion of Okun that aside from obtaining a greater force for the solution of oxygen, the only improvement was a considerably greater sludge density in the high purity oxygen unit. Although the oxygenated sludge may settle better, its activity or ability to assimulate the waste per gram of organisms appeared to be the same as that for air.

In 1955, the first large scale field study was undertaken to determine the advantages of the process described above (9). No direct comparison was made with air aeration; however, the possible reduction in reactor volume when using oxygen aeration was estimated to be thirty percent, because a higher MLSS could be maintained. This estimate was considerably smaller than the savings in reactor volume expected from the initial laboratory study, and was probably the result of the lower MLSS maintained in the field study.

Batavia

The oxygen system evaluated at Batavia, New York, was the first large-scale design of the UNOX process (10). The oxygen unit used in this study differed in several ways from the earlier Okun design. A typical three-stage UNOX reactor is shown schematically in Figure 3. Oxygen was transferred within the reactor rather than by an external absorber. Oxygen was fed into the first section under

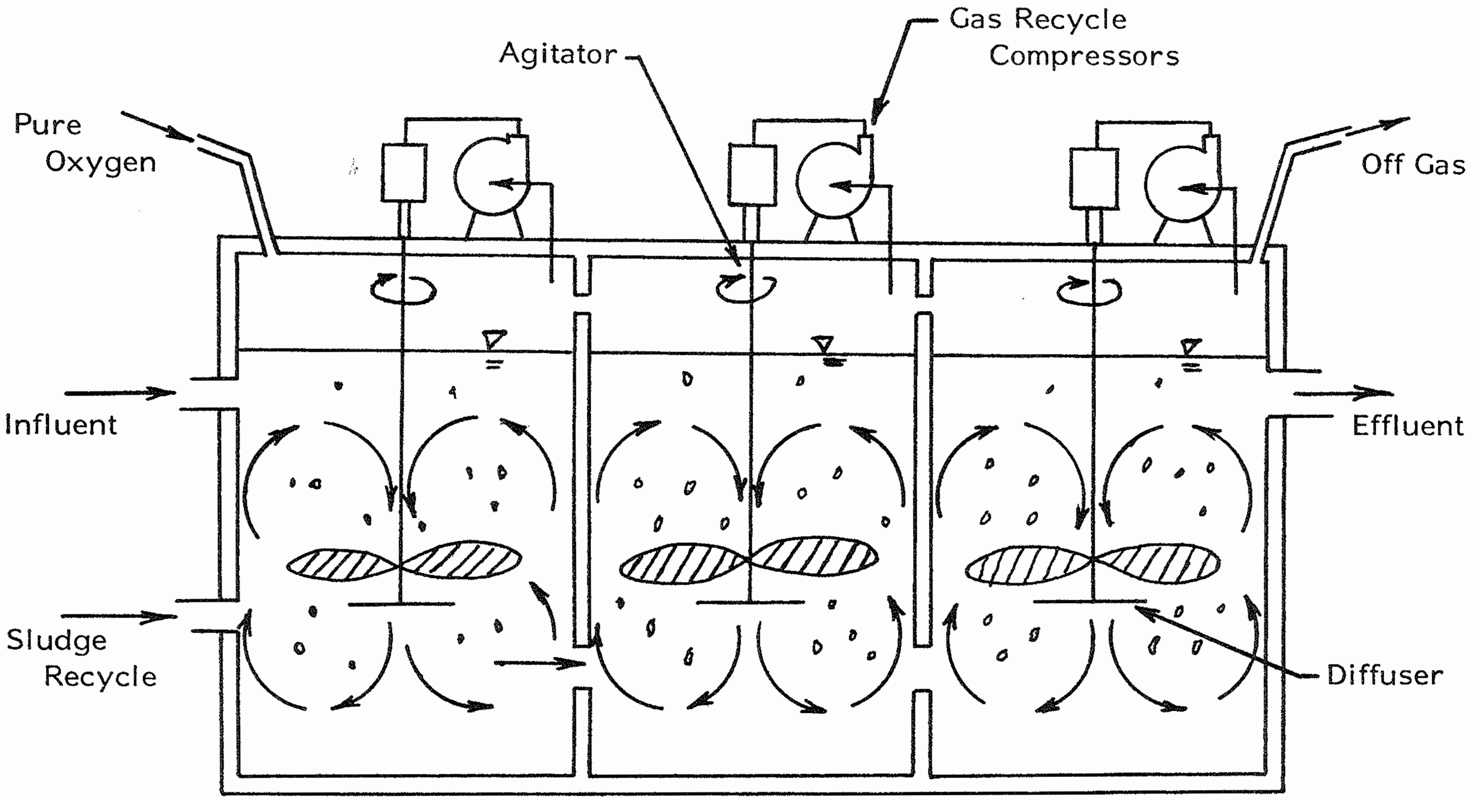

FIGURE 3. UNOX PROCESS USED IN FULL SCALE STUDIES AT BATAVIA, N.Y.

a pressure of several inches of water and allowed to pass in series from compartment to compartment. Air was excluded and oxygen prevented from escape by tight-fitting covers over each compartment. The last section allowed off-gas to vent under automatic pressure control. Each compartment was equipped with a high speed turbine mixer to mix the contents and to help shear the fine oxygen bubbles released by the diffuser located below the turbine. Some oxygen transfer occurred at the gas-liquid interface near the top of the compartment, but the majority of the transfer was accomplished via gas recycle compression and diffusion. The recycle compressor would pump oxygen-enriched gas from the top of each compartment to the diffuser for the purpose of utilizing as much of the oxygen as possible. The compartments were in series and a separate oxygen compressor and diffuser installed in each. As a result, the transfer efficiency was improved over the Okun system to about 90 or 95 percent.

Tests were conducted under three modes of operation. Phases I and III were designed to provide side-by-side comparison of air and oxygenated systems. The reactor volumes and feed rates were nearly the same for the parallel systems. In phase II, only the oxygenated system was operated. Results of the study are summarized in Table I. Some of the data are averages listed in or calculated from the data in the report. A review of the data indicates

TABLE I.

SUMMARY OF BATAVIA DATA

	Phase I		Phase II	Phase III	
	Air	Oxygen	Oxygen	Air	Oxygen
Influent Flow, MGD	1.97	1.91	2.53	1.29	1.44
Influent BOD , mg/l	159	159	220	262	262
Influent VSS, mg/l	152	152	123	332	332
Percent BOD Removal	90	92	90	88	94
Aeration Detention Time, hrs.	4.0	4.1	1.5	3.0	2.8
MLVSS, mg/l	1740	2210	4450	2580	4310
Mean Sludge Age, days	2.4	5.4	3.1	1.5	14.5

that higher mixed liquor suspended solids concentrations could be maintained in the oxygenated system. The greater ease of oxygen transfer and the possibility of a more rapidly settling sludge in the oxygenated system, account for this improvement.

The sludge yield for the Batavia plant was computed to be 1.38 lb VSS/lb of BOD_5. The resulting yield coefficient was significantly greater than any reported in other studies (11). These units received a feed containing high concentrations of suspended solids resulting from no primary sedimentation. One pound of BOD results in about 0.5 pound of biomass and another 0.5 pound of solids in the influent appeared as sludge yield. Therefore, the actual sludge productivity or yield is estimated at 0.4 lb to 0.6 lb VSS/lb BOD_5.

Newtown Creek

A second-generation multi-stage UNOX process in which the recirculating compressors and rotating spargers are replaced with submerged aerators or mixers is installed at Newtown Creek Plant in New York City (12). Oxygen transfer is accomplished by gas entrainment and solution. Performance of the secondary system, average sludge characteristics, and average loadings are summarized in Table II. Process performance to December 1972 has confirmed the high rate capabilities first seen at Batavia. The sludge productivity was 0.51 lb VSS/lb BOD_5 and the oxygen utilization was 96 percent.

TABLE II.

SUMMARY OF NEWTOWN CREEK PERFORMANCE DATA AND LOADINGS

Total BOD_5 Removal Efficiency (percent) = 95

Soluble BOD_5 Removal Efficiency (percent) = 96

Total COD Removal Efficiency (percent) = 83

Suspended Solids Removal Efficiency (percent) = 91

MLSS (mg/l) = 4760

MLVSS (mg/l) = 3960

SVI (ml/gm) = 42

Sludge Retention Time, SRT
(lb MLVSS/lb VSS wasted-day) = 3.07

F/M Loading (lb BOD_5 applied/lb MLVSS/day) = 0.67

F/M Loading (lb COD applied/lb MLVSS/day) = 1.58

Organic Loading (lb BOD_5 applied/1000 cu ft-day) = 148

Organic Loading (lb COD applied/1000 cu ft-day) = 389

Clarifier Overflow Rate (gpd/sq ft) = 977

Clarifier Mass Loading (lb solids/sq ft-day) = 50.7

Clarifier Weir Loading (gpd/ft) = 133000

Brewery Waste - 1970

A bench-scale study was conducted at a large brewery to determine possible advantages offered by high purity oxygen treatment (13). Three levels of dissolved oxygen were investigated, including about one mg/l (air), seven mg/l (air plus oxygen) and 15 mg/l (oxygen). The growth kinetics were the same for all levels, but the sludge settled better and appeared to show higher organic removal rates under oxygen conditions. Unfortunately the reported data were not complete and an independent analysis could not be performed.

One of the most interesting points of the study was the relationship between sludge volume index and the food to microorganism ratio, shown in Figure 4. The SVI appeared to increase with the larger values of F/M during air and air plus oxygen runs to a maximum SVI of 1000 at the highest F/M ratio for the air plus oxygen system. At this point, the oxygen system was started and the SVI averaged about 50 regardless of the F/M ratio. The cell yield developed in the brewery study was 0.56 lb VSS/lb BOD removed.

Blue Plains

A multi-stage covered-tank oxygenation pilot system of Union Carbide design (Figure 5) has operated continuously since June, 1970 at the Blue Plains Pilot Plant (14). The average system F/M loadings are 0.27 to 0.50 lb BOD applied/day/lb MLVSS under aeration

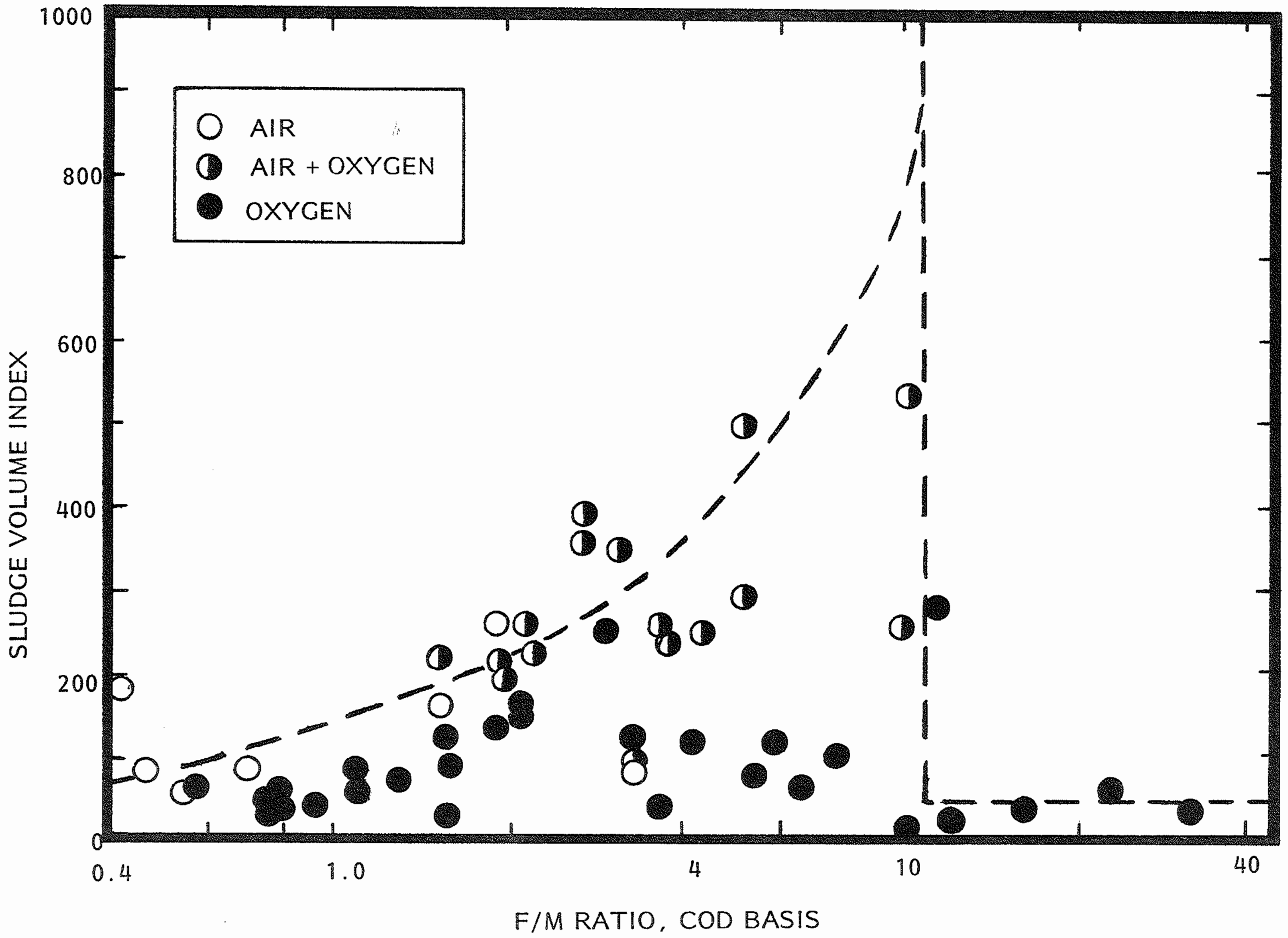

FIGURE 4. RELATIONSHIP OF SVI TO F/M RATIO FROM BREWERY WASTE

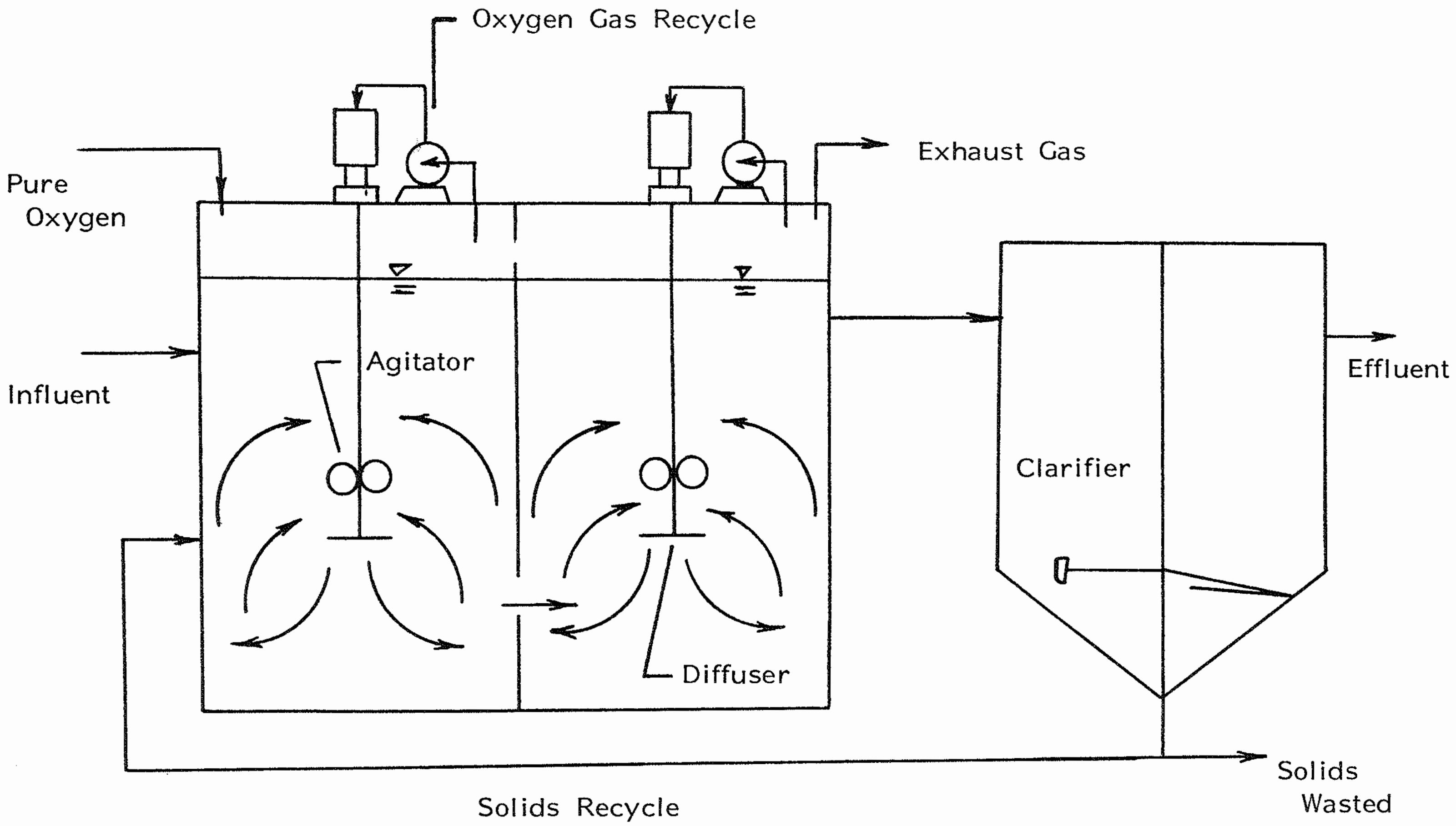

FIGURE 5. SCHEMATIC DIAGRAM OF BLUE PLAINS SYSTEM

at a sludge retention time (SRT) of more than five days. When the system has operated at an SRT of less than five days, F/M loadings have risen to as high as 1.0 lb BOD_5/day/lb MLVSS. Corresponding average volumetric organic loadings at a SRT above five days have been 57 to 185 lb BOD applied/day/1000 ft of aerator volume. The hydraulic detention times varied from 1.5 to 2.8 hours. For all loadings investigated, virtually complete oxidation of soluble BOD has been observed. The average effluent soluble BOD concentration is 2 to 3 mg/l but never greater than 5 mg/l. Total BOD and suspended solids removal are a function of clarifier performance. Effluent COD and TOC concentrations typically have ranged from 35 to 60 and 15 to 20 mg/l, respectively. The sludge yield is 0.38 to 0.47 lb VSS/lb BOD removed.

Results of the long-term Blue Plains study illustrate that the design of high purity oxygen systems should be an integrated package consisting of a biological reactor, a clarifier, and sludge handling facilities. The design should consider the worst anticipated climatic conditions at a given site. Clarifier sizing should be specifically tailored to the design and anticipated operating conditions of the reactor. The two basic methods of achieving a desired F/M loading are: (a) controlling the size of the reactor and/or (b) controlling the MLSS concentration. For example a unit employing a small reactor and high MLSS can have the same F/M loading as a large reactor and low MLSS. Selection of a small reactor and a small clarifier cannot sucessfully be coupled in a design unless a greatly reduced MLSS concentration is maintained. However, opting for this selection will increase F/M loading, biological sludge production, and required sludge handling capacity and costs.

SIMPLOX

A single-stage, covered-tank oxygenation system, SIMPLOX, is shown schematically in Figure 6 (12) and incorporates an inflated dome-type cover to contain the oxygen-rich atmosphere over the aerator. This approach is intended primarily for upgrading existing air-activated sludge plants by utilizing conventional air blowers and coarse-bubble air diffusers to recirculate oxygen gas. Oxygen gas is introduced to the aerator through a fine bubble sparger on the tank bottom and on the opposite side wall from the conventional air diffusers. Power required for oxygen solution is greater for the SIMPLOX process than for the multi-stage systems because: (a) the equipment used for transfering oxygen is modified air aeration equipment and not specifically tailored for oxygen, and (b) the gas phase above the mixed liquor is completely mixed and assumes the same oxygen composition as the exhaust gas stream. Therefore, the driving force for dissolving oxygen in wastewater is less than in the first stages of multi-stage aerators. However, capital costs for converting an existing aeration tank from air to oxygen service should be significantly less with the SIMPLOX approach

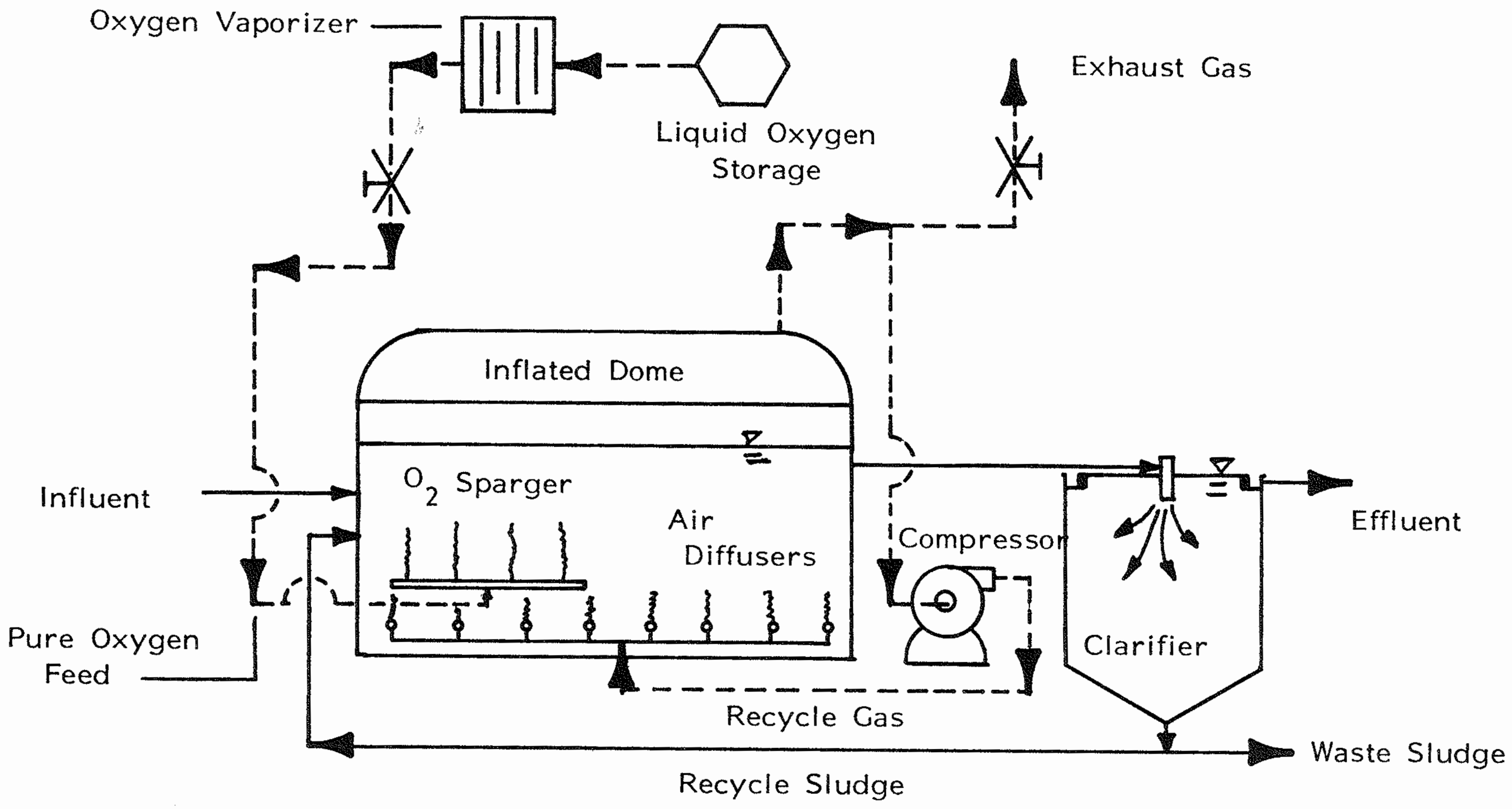

FIGURE 6. SCHEMATIC DIAGRAM OF DIFFUSED AERATION SYSTEM MODIFIED TO OXYGEN GAS, SIMPLOX SYSTEM

because staging baffles and multiple oxygen solution equipment are not required. Any of several activated sludge flow configurations, including plug flow, complete mix, and step aeration, can be used as desired. Since the gas phase is completely mixed, exhaust oxygen, carbon dioxide, and inert gases can be bled from any point of the inflated dome.

Testing is being evaluated under four modes of operation divided into phase I, II, III, and IV. Phases I to IV will alter the aeration detention times by varying the percent of the oxygen aerator in use and/or the flow rate. As of December 1972 only the ultra conservative Phase I had been completed. Substrate removals, sludge characteristics, and average system loadings are summarized in Table III. The sludge yield was 0.23 lb VSS/lb BOD removed. The inflated dome was determined to be unsuitable for permanent installation because of leaks or punctures in the polyvinyl material.

Future Research and Development Projects

Utilization of oxygen aeration for activated sludge treatment will be evaluated in two new projects which are approaching the development and experimental application stages (December 1972). One project is the Martin Marietta, while the second is Downflow Bubble Contact Aeration.

The Martin Marietta Project is a development of a unique

TABLE III.

SUMMARY OF SIMPLOX, LAS VIRGENES, CA., PERFORMANCE DATA AND LOADINGS

Total BOD_5 Removal Efficiency (percent)	=	97
Total COD Removal Efficiency (percent)	=	77
Soluble COD Removal Efficiency (percent)	=	72
Suspended Solids Removal Efficiency (percent)	=	88
MLSS (mg/l)	=	3700
MLVSS (mg/l)	=	2950
SVI (ml/gm)	=	99
Sludge Retention Time, SRT (lb MLVSS/lb VSS wasted-day)	=	78
F/M Loading (lb BOD_5 applied/lb MLVSS/day)	=	0.07
F/M Loading (lb COD applied/lb MLVSS/day)	=	0.13
Organic Loading (lb BOD_5 applied/1000 cu ft-day)	=	13
Organic Loading (lb COD applied/1000 cu ft-day)	=	24
Clarifier Overflow Rate (gpd/sq ft)	=	417
Clarifier Mass Loading (lb solids/sq ft-day)	=	16.7

fine bubble diffuser (12). One of the many potential applications for the diffuser is in an open-tank oxygen activated sludge process. The Martin Marietta Corporation claims that: (a) uniform oxygen bubbles of less than 0.2 mm diameter are developed by passing both oxygen gas at 30 psi and water or sewage through the device and out through slot-type openings, and (b) these bubbles when discharged at the bottom of a vessel with five feet of water depth are completely dissolved before reaching the surface.

Major points of research interest in the project are: (a) oxygen utilization efficiency in an open-tank, (b) mixed liquor recirculation rates and power requirements, (c) diffuser self-cleansing (non-clogging) capabilities, and (d) shearing effect, if any, on mixed liquor particles from continuous recirculation through the pumps and diffusers. In the event that floc disruption does occur, a short detention biological reflocculation tank (gentle mixing, no chemicals) will be interposed between the reactor and the clarifier. Two aspects of system design which cannot be defined adequately because of the scale of this plant are diffuser mixing characteristics and possible additional mixing requirements for the large aeration tanks.

A simple system has been developed by Speece et al. (15) for efficient absorption of oxygen and compressed air into water. It has been descriptively named, Downflow Bubble Contact Aeration (DBCA). DBCA provides for the passage of water or sewage downward through an open-bottomed, expanding cross section hood or cone. The water enters at the highest velocity at the top and leaves at the lowest velocity from the bottom. The inlet velocity is designed to be greater than the buoyant velocity of the bubbles. The exit velocity is designed to be less than the buoyant velocity of the bubbles. Thus the bubbles are trapped inside the hood as the water flows through it.

The hood can fill with bubbles if the ratio of bubble interfacial area to water volume is high. Also, there is considerable turbulence, resulting in high gas transfer rates. Bubbles injected into a 10-ft deep tank of water rise to the surface and escape in about 10 seconds or less. By comparison, bubbles can be maintained in contact with water in DBCA for an indefinite period. Thus, efficient transfer of oxygen is possible because of the prolonged contact times.

Results of studies incorporating the DBCA indicates oxygen absorption efficiency of 80 to 90 percent at an oxygen ratio of 0.5 percent of the water flowrate by volume. At sea level and a temperature of 68 degrees Fahrenheit, an increase of 5.7 to 6.4 mg/l of dissolved oxygen in water passing through the system would result. A number of potential applications of DBCA are being investigated.

Kinetics and Settleability

Ball and Humenick (16, 17) noted that there was no significant difference in the sludge yield or substrate removal kinetics under oxygen and air. No significant difference in the initial settling velocities of the activated sludge developed under air or high purity oxygen was observed when compared at the same suspended solids concentrations, sludge age, and mixing intensity. Initial settling velocities were a function of mixing intensity. Because the increased rate of oxygen transfer under oxygenation allowed reduced gas mixing in the reactor, the lower mixing levels resulted in improved settling velocities. Typical mixing intensities under oxygenation averaged 40 ft/sec-ft compared with 100 to 160 ft/sec-ft for full-scale aeration plants. Full-scale settling studies were used to develop a method for determining practical limits for sludge underflow concentrations and sludge recycle rates. A number of sludge characterization tests also were reported, including dehydrogenase activity, extracellular carbohydrates, cellular bound water and microscopic examinations, with no apparent differences found between oxygenated and aerated sludges.

Results of the study showed that high purity oxygen was a realistic alternative to conventional aeration with additional costs for producing and distributing the oxygen economically justified in some cases. The benefits are primarily related to an ability to carry increased MLSS concentrations in the reactor rather than from significant fundamental changes to the organisms. Higher MLSS concentrations can result in decreased capital expenditures for the reactor and an increase in the sludge age. An increase in sludge age will result in a reduction in capital and operating expense as a result of the decrease in net sludge to be wasted.

Ball (17) operated two integrated biological-chemical wastewater treatment systems in parallel using settled municipal wastewater as feed. The aeration source for the biological reactors was conventional low-pressure air in one system and for the other system high purity oxygen with additional mixing was provided.

No significant differences in kinetic constants were apparent. The yield was 1.0 lb MLVSS/lb BOD removed and the decay coefficient was 0.4 lb MLVSS destroyed/lb MLVSS. The maximum rate of substrate utilization was 12.5 lb BOD_5/day-lb MLVSS. BOD removal exceeded 90 percent at substrate removal rates up to 1.36 lb BOD_5/lb MLVSS-day.

Settling tests indicated that no significant difference existed between sludges developed under aeration or oxygenation for the same mixing levels. At low mixing levels, possible under oxygenation, significantly higher solids flux rates were achieved for the same MLSS concentration than in the aerated system.

Oxygen Generation

A number of physical or chemical processes are available for the production of oxygen. Among the most common are electrolysis, adsorption, and low temperature distillation of air. Presently, there are two methods of large-scale oxygen production for wastewater treatment: (a) a low temperature distillation process (Cryogenic) and (b) an adsorption process (Pressure Swing Adsorption).

The cryogenic process extracts oxygen from the air by means of liquefaction/distillation (30). To liquefy, air must be cooled to a temperature of minus 300 degrees Fahrenheit. Distillation of the liquefied air occurs in two distillation columns that take advantage of the different boiling points of the oxygen, nitrogen and the other atmospheric components. However, the carbon dioxide and water vapor must be removed from the process air stream through another extraction before entry into the distillation columns. The normal oxygen purity of cryogenic generators is in excess of 95 percent, although the low pressure process can achieve 99.5 percent purity if desired.

The Pressure Swing Adsorption process can generate up to 95 percent pure oxygen from air. The system involves adsorption at a high pressure and desorption at a low pressure (31). Compressed air is introduced into a fixed bed of adsorbent in which nitrogen is preferentially adsorbed to produce oxygen-rich gas. When the bed approaches saturation, the bed pressure is reduced to desorb nitrogen and regenerate the bed. A purge by some of the product or an intermediate process stream is used to increase the regeneration efficiency. Generally, two or more adsorbent beds are employed to insure continuous oxygen generation. One bed is in the adsorption phase and the other bed is in the regeneration phase. Many pressure swing adsorption processes are available and the major differences in these processes are: (a) the type of adsorbent used, (b) the number of beds, (c) operating pressures, and (d) operating sequence. The advantages claimed for pressure swing adsorption are lower cost, easy operation, quick start up, and flexibility in production capacity as well as oxygen concentration.

Typical cost curves for generating oxygen cryogenically and by pressure adsorption systems are presented in Figure 7. These data indicate that at oxygen generation rates less than 80 tons per day the cost per ton of oxygen is less for the PSA system. The cost of oxygen generation is between $8 and $18/ton.

BIOLOGICAL KINETICS

Two principal relationships of biological kinetics are the equations relating cell growth and decay rates to substrate removal. The two relationships will be derived from the model shown in Figure 8.

The relation of biological growth and endogenous metabolism to the rate of removal of substrate may be derived by a materials balance of biological solids:

$$dX/dt = X_o - X_e - X_w + Y(S_o - S_e) - k_d X \quad . \quad . \quad . \quad . \quad 1$$

where:

dX/dt = Change in amount of cells (VSS) per unit time, lbs/day

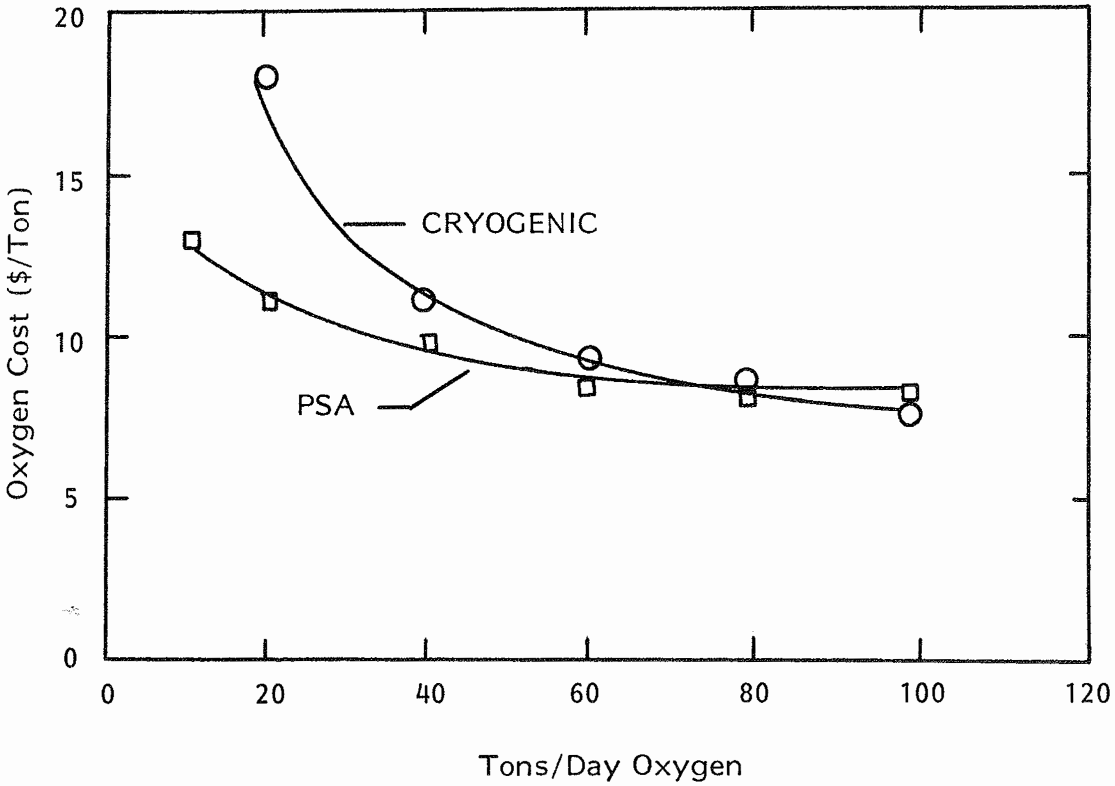

FIGURE 7. COST OF OXYGEN GENERATION

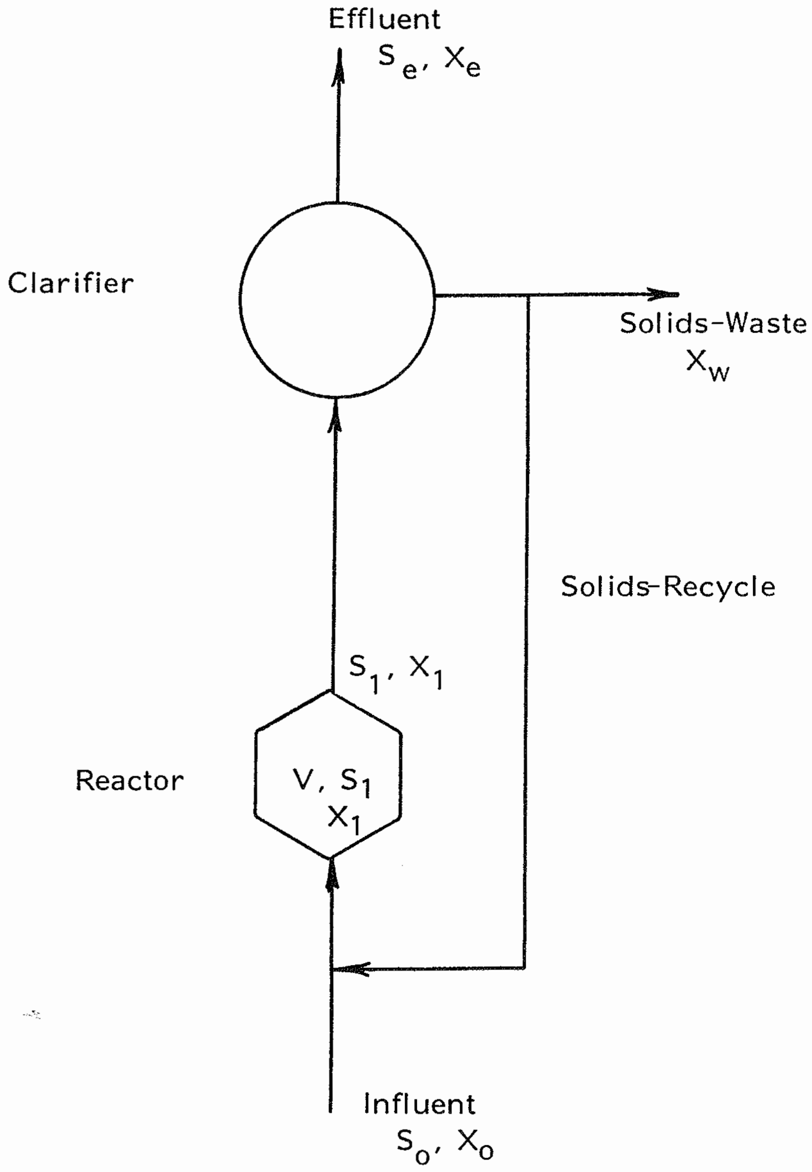

FIGURE 8. SCHEMATIC DIAGRAM FOR BIOLOGICAL KINETICS

X_o = Amount of cells (VSS) in influent, lb/day

X_e = Amount of cells (VSS) in effluent, lb/day

X_w = Amount of cells (VSS) in waste stream, lb/day

X = Amount of cells (VSS) synthesized/substrate removed, lb/lb

S_o = Total substrate in influent, lb/day

S_e = Total substrate in effluent, lb/day

k_d = Rate of endogenous respiration, lb/lb/day

In practice, the major difficulty with Equation 1 is that the Volatile Suspended Solids (VSS) content is assumed to estimate the actual amount of cells in the system. If the system influent has appreciable VSS, the yield coefficient, Y, will be higher than the true value of VSS synthesized per unit of substrate removed, as there is no way to distinguish between VSS which exist as a result of substrate utilization and the VSS which are introduced to the system by the influent.

In spite of the shortcomings of using VSS as the measure of active biological mass, the problems associated with other methods of determining the active mass are even more complex, in addition to the fact that insufficient data are available to establish confidence in some of the other methods (18).

Kinetic design parameters are usually determined in laboratory studies during normal technical investigations. If a completely soluble influent is used and the reactor is completely mixed with no recycle or wasting at steady-state conditions, Equation 1 can be simplified as follows:

$$X_e = Y(S_o - S_e) - k_d X \quad . \; . \; . \; . \; . \; . \; . \; . \; . \; . \; . \quad 2$$

The second relationship is developed from the empirical equation developed for microbiological cultures (19):

$$\mu = \hat{\mu} S/(K_s + S) \quad . \; . \; . \; . \; . \; . \; . \; . \; . \; . \quad 3$$

where:

μ = specific growth rate, lbs/lbs/day

$\hat{\mu}$ = maximum specific growth rate, lbs/lbs/day

K_s = substrate concentration at $\frac{1}{2}\hat{\mu}$, mg/l

By assuming the relationship:

$$\mu = Yq \text{ and } \hat{\mu} = Y\hat{q}$$

where:

q = rate of substrate removal, lbs/lbs/day

$\hat{q}$ = maximum rate of substrate removal, lbs/lbs/day

the following useful relation is obtained:

$$q = \hat{q} S/(K_s + S) \quad . \; . \; . \; . \; . \; . \; . \; . \; . \; . \; . \quad 4$$

The reciprocal expansion of Equation 4 can then be plotted by obtaining values of the substrate removal rate (q) and the substrate concentration (S), and the equation constants determined.

If Equation 2 is divided by the VSS in the system (X), it can be written in the more familiar form:

$$\theta_c^{-1} = Yq - k_d \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \qquad 5$$

where:

θ_c = mean solids retention time, day = X/X_e

$q = (S_o - S_e)/X$, lbs/lbs/day

The constants for Equation 5 can be obtained by a plot of the rate of substrate removal (q) versus the inverse of the mean solids retention time.

Therefore, a rational design of a biological process can be made for that particular type of wastewater with the information developed from the plots of Equations 4 and 5 (20).

SLUDGE SETTLING

The separation of the solids in the mixed liquor for recycle from the treated wastewater is probably the most critical step in the activated sludge process. The degree of separation depends on the amount of dispersed cells that remain in the supernatant plus the amount of sludge which would settle, but is carried over into the effluent because of poor clarifier design or uncontrollable hydraulic variations. The amount of dispersed cells and the rate of subsidence of the settleable fraction will vary from day to day. The importance of settleability in design has resulted in the development of several measures for comparison.

The sludge volume index, SVI, introduced in 1934 is the most commonly accepted measure of settleability (21). Sludge volume index is defined as the volume, in milliliters, occupied by one gram of activated sludge after settling in a one liter graduate cylinder for 30 minutes (22). Data, obtained by recording the solids interface with respect to time, can then be subjected to a variety of manipulations intended to model the actual unit operation of clarification or thickening to produce a design for that sludge.

Dick and Vesilind (23) conducted in-depth studies of the SVI test and concluded that two sludges could have the same index value although grossly different in settling characteristics as shown in Figure 9. Some of the factors influencing rate of subsidence are temperature, cylinder depth, cylinder diameter, sludge depth or initial concentration, and mixing history (24).

Another factor which affects settleability is the mixing level in the reactor before clarification. It is important, therefore, to calculate the mixing level of the sludge under investigation in order to compare the results of several tests. The formula for calculating the mixing level as determined by Parker _et al_. (26) follows:

For bubble mixing:

$$G = Q'hg/v \quad . \; . \; . \; . \; . \; . \; . \; . \; . \; . \; . \; . \; . \quad 6$$

where:

G = Mixing level in ft/sec-ft

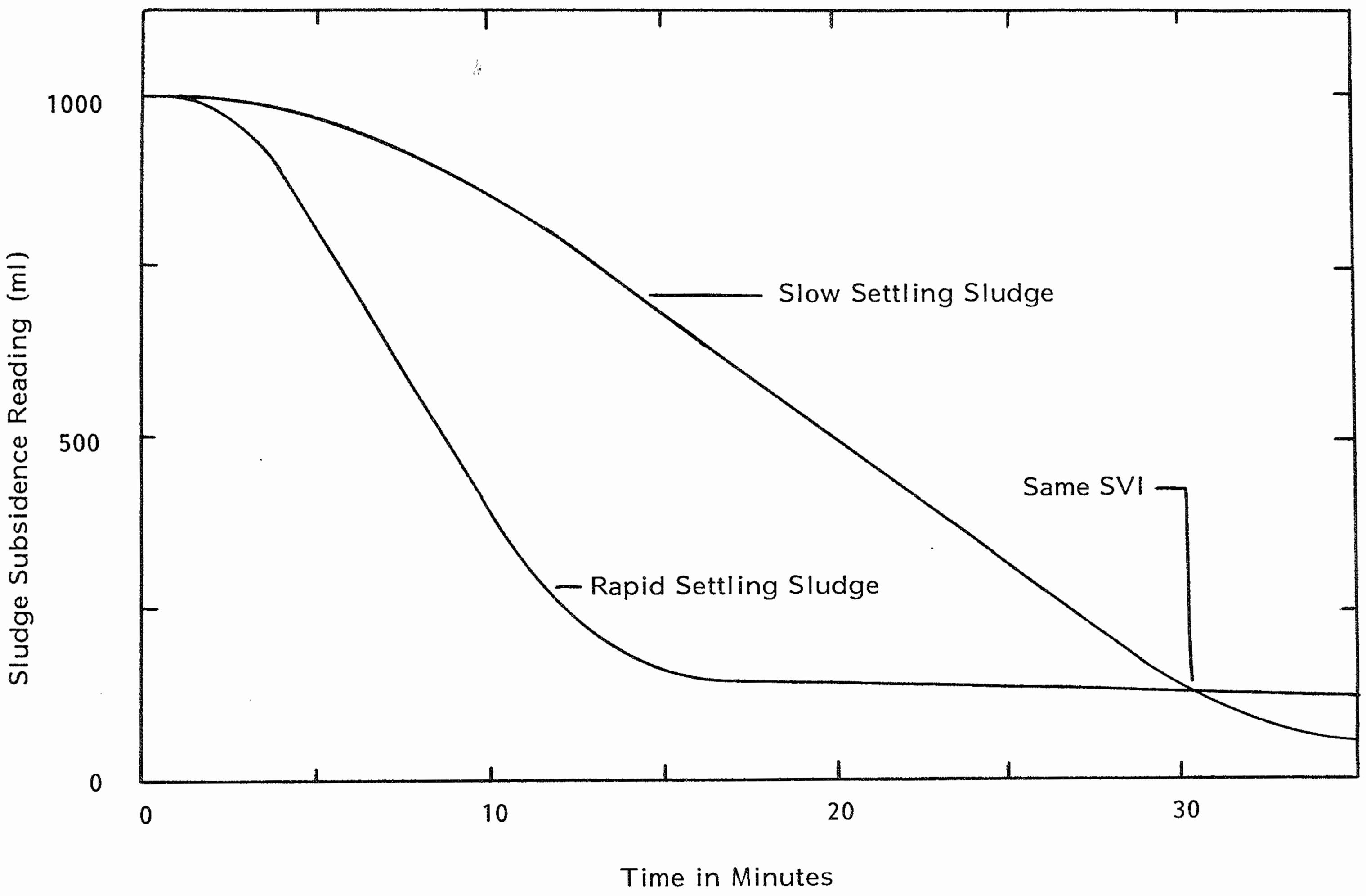

FIGURE 9. ILLUSTRATION OF TWO SLUDGES WITH SAME SVI BUT DIFFERENT SETTLING CHARACTERISTICS

Q' = Gas flow in l/min-unit volume

h = Diffuser depth, inches

g = 32.2 ft/sec

v = Kinematic viscosity of water, sec /ft

There are two conditions which result in poor settling of activated sludge, namely dispersed growth and sludge bulking. Dispersed growth is typical in systems operating at low sludge ages. Tenny and Stumm (27) believed that poor settling occurs at low mean cell retention times because natural polymers, such as complex polysaccharides and polyamino acids, are not excreted at a rate sufficient to provide a proper polyelectrolyte "dose" for flocculation. These studies of flocculating dispersed growth with synthetic polymers have supported the idea of cell self-polymerization.

Sludge bulking has plagued the treatment plant operator since the introduction of the activated sludge process (28). Bulking occurs when filamentous microorganisms constitute a significant percentage of the total biomass, resulting in a marked decrease in sludge settling velocity in the clarifier and a relatively high SVI (29).

SUMMARY OF LITERATURE REVIEW

The use of high purity oxygen in the activated sludge process is receiving increasing attention in wastewater treatment plant design and construction. The concept is more than 20 years old, but has received serious consideration only during the last 4 or 5 years with the development of several cost-effective systems for dissolving and utilizing oxygen in an aeration tank.

The use of oxygen opens up the possibility of higher MLSS concentrations within the aeration basin while completely satisfying the correspondingly high oxygen demand rates. Mixing requirements for MLSS concentrations between 5000 and 10,000 mg/l in oxygen activated sludge systems are not similar to lower MLSS concentrations in air activated sludge systems. A higher sludge density can be realized and the net sludge yield can be reduced while net sludge age increases.

The kinetic behavior of biological processes may be satisfactorily described by two basic relationships: (a) biological growth and (b) substrate concentration biomass. The mode of aeration does not affect the use of these kinetic relationships.

Activated sludge settleability varies with the MLSS and mixing level before sedimentation. The design of clarifiers requires the construction of a solids transport curve for a given sludge. A rational formula for bubble mixing is available to calculate mixing levels.

IV. MATERIALS AND METHODS

The Downflow Bubble Contact Aeration System was located at Govalle Wastewater Treatment Plant, Austin, Texas. The unit was set up to treat the effluent from primary clarifiers and existing equipment was utilized.

EXPERIMENTAL EQUIPMENT AND SYSTEMS

Downflow Bubble Contact Aerator

Efficient absorption of oxygen and/or air into water has been developed and demonstrated by Speece, et al. (15). Downflow Bubble Contact Aeration (DBCA) provides for the passage of water downward through an open-bottom expanding cross section hood, Figure 10. The water enters at its maximum velocity (4 to 8 ft/sec) and reaches a minimum velocity (0.1 to 0.2 ft/sec) at the widest point of the hood. The inlet velocity is designed to be greater than the buoyant velocity of the bubbles. The exit velocity is designed to be less than the bouyant velocity of the bubbles. Hence, the bubbles are trapped inside the hood.

The hood fills with bubbles and provides a high ratio of bubble interfacial area to water volume. Considerable turbulence is present and increases the gas transfer. In comparison with normal

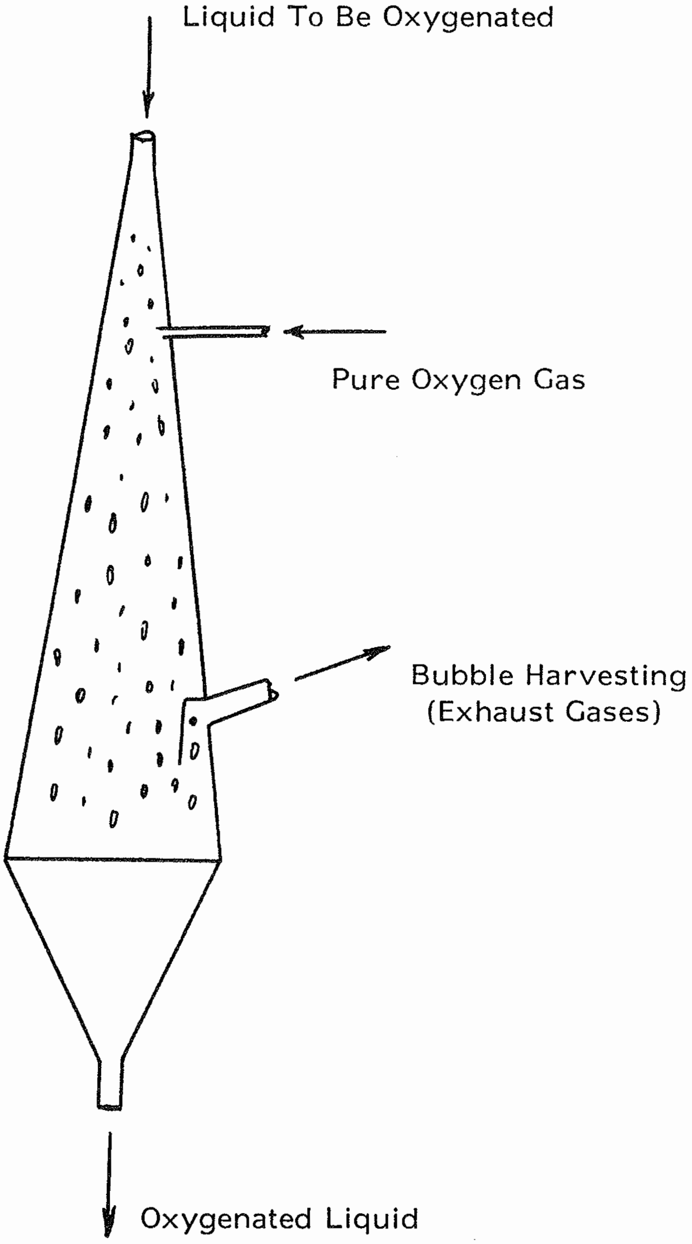

FIGURE 10. DOWNFLOW BUBBLE CONTACT AERATOR

transfer (bubbles rising in a 10 ft tank with a contact of approximately 10 seconds), DBCA gives gas-water contact for prolonged periods of time.

Nitrogen stripping and oxygen absorption occur simultaneously in the DBCA resulting in a change of gas composition of the bubbles from approximately 100 percent oxygen to approximately 100 percent nitrogen and carbon dioxide before harvesting. The bubbles would be completely absorbed and disappear if the water contained no dissolved gases. Transfer efficiencies usually range from 80 to 95 percent depending upon the oxygen-water ratio.

DBCA Activated Sludge System, Unit I

The initial continuous flow activated sludge system incorporating DBCA as the transfer unit is represented by Figure 11. The sludge was recycled and passed through the DBCA. The sludge is being oxygenated in the units presented in Figures 11 and 12. The raw influent entered the reaction portion of the vessel and immediately became mixed with freshly oxygenated sludge. The large tank was operated as a reactor-clarifier, the lower portion the reactor and the upper portion the clarifier.

To aid sludge separation and/or mixing, two modifications of the reactor-clarifier system were utilized. Initially, an artificial media mat was incorporated just above the sludge recycle line in

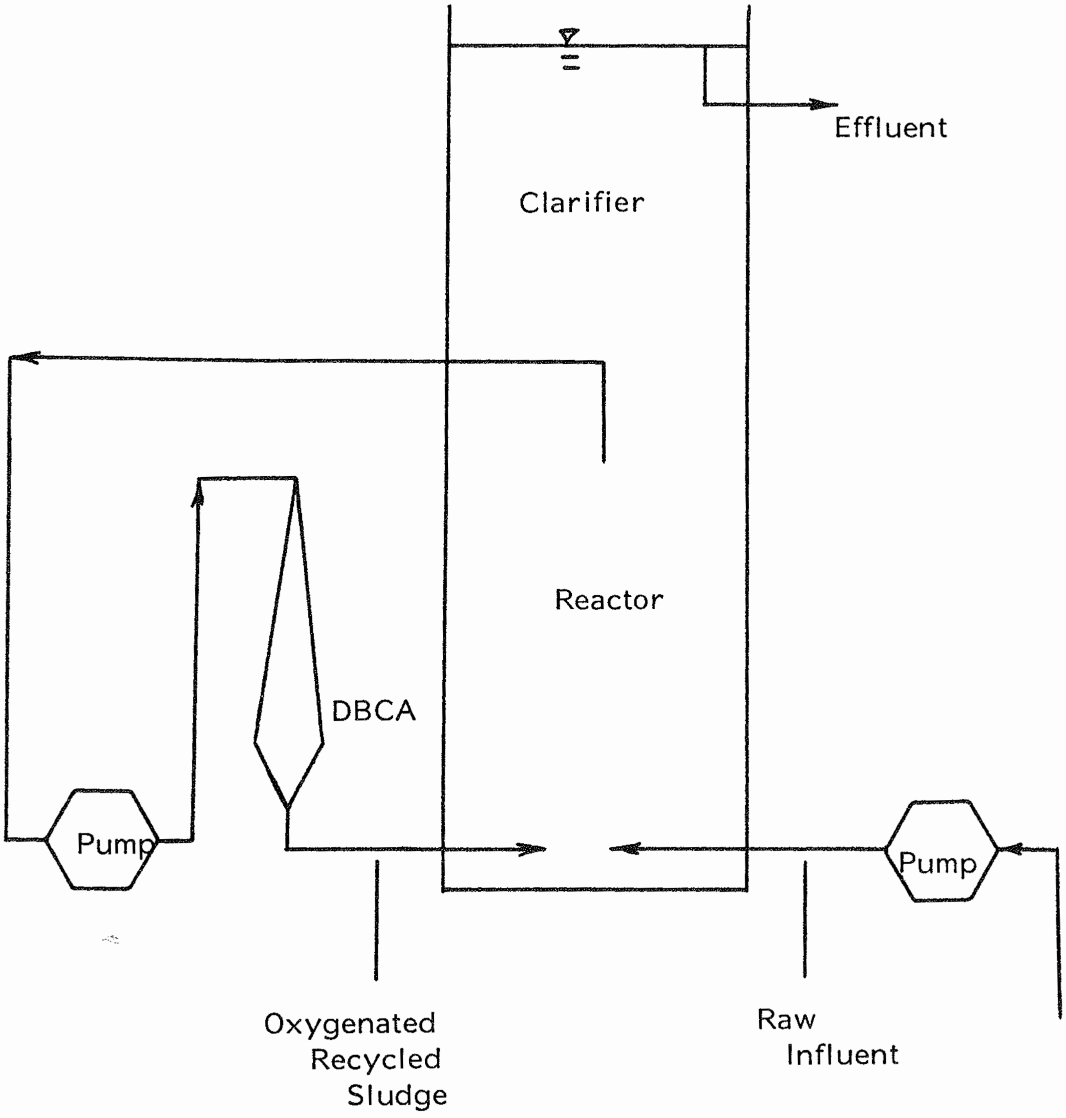

FIGURE 11. DBCA ACTIVATED SLUDGE SYSTEM, UNIT I

in order to shear oxygen bubbles from the sludge physically. Secondly, mixing was intensified in hoped of shearing off the minute oxygen bubbles. Mixing was increased by the controlled release of compressed air at strategic locations within the reactor.

In all cases, the operation parameters and loadings were essentially the same. The hydraulic overflow rate on the clarifier was 710 gpd sq ft. The food to microorganism rate, F/M was 0.3 lb BOD_5/lb MLSS-day, and oxygen was supplied at the rate of one lb O_2/lb BOD_5. The design flow of the system was 3.5 gpm.

DBCA Activated Sludge System, Unit II

The original reactor-clarifier system was divided into three distinct units, Figure 12, to enhance the sludge separation characteristics. The transfer unit was the DBCA, however, a separate biological reactor and clarifier were used. The system was operated in two modes: with and without mechanical mixing to increase shear.

In both cases, the operational parameters and loading rates were the same. The hydraulic overflow rate was maintained at 920 gpd/sq ft. The food to microorganism ratio, F/M was 0.1 lb BOD_5/lb MLSS-day. The oxygen flow rate was adjusted to maintain approximately one lb O_2/lb BOD_5. The flow rate in the system was two gpm.

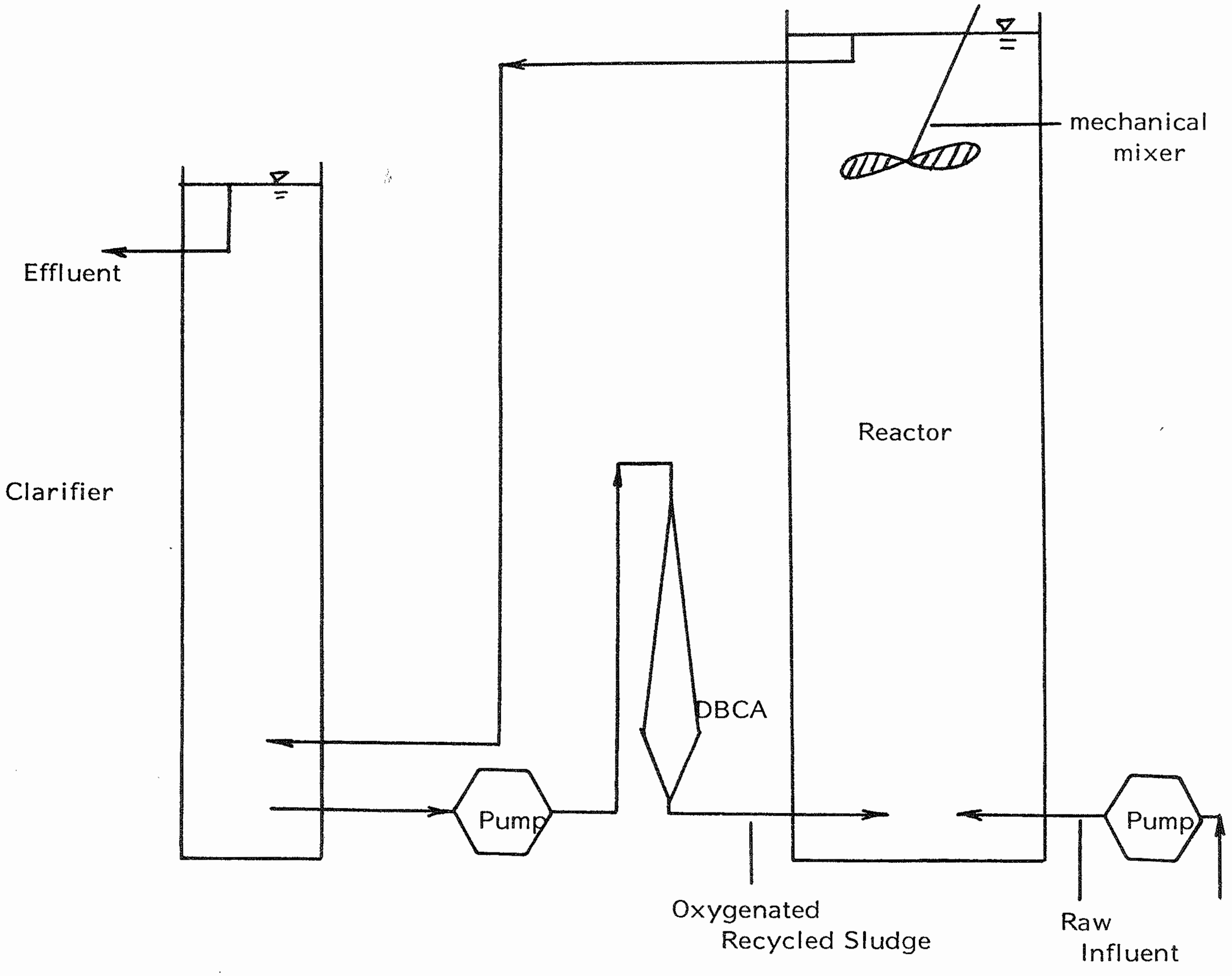

FIGURE 12. DBCA ACTIVATED SLUDGE SYSTEM, UNIT II

DBCA Activated Sludge System, Unit III

Unit III is a complete change in design. The oxygen transfer section (DBCA) was used to oxygenate the raw influent and not the biological sludge mass. Three distince separate units were employed; namely: transfer, reactor, and clarifier, Figure 13.

The sludge was recycled and mixed in the lower portion of the reactor with the oxygenated raw influent. The recirculation ratio was 1 : 1. Mixers were placed in the upper portion of the reactor to aid in shearing oxygen bubbles from the sludge.

The hydraulic overflow rate was 410 gpd/sq ft which was dictated by the original design of DBCA, Unit I. The oxygen flow was maintained at one lb O_2/lb BOD_5. The food to microorganism ratio, F/M, was 0.45 lb BOD_5/lb MLSS. The flow rate through the system was two gpm.

Oxygen Metering and Regulation

Oxygen was purchased in standard 244 cu ft cylinders. Pressure was regulated to 100 psi by a two-stage regulator to control the gas flow. In this manner, gas flow was independent of operating pressures.

Calibrated and correlated flow meters were utilized to control the flow with corrections for the type of gas and the pressure. Determination of the new calibration curve was accomplished by a wet

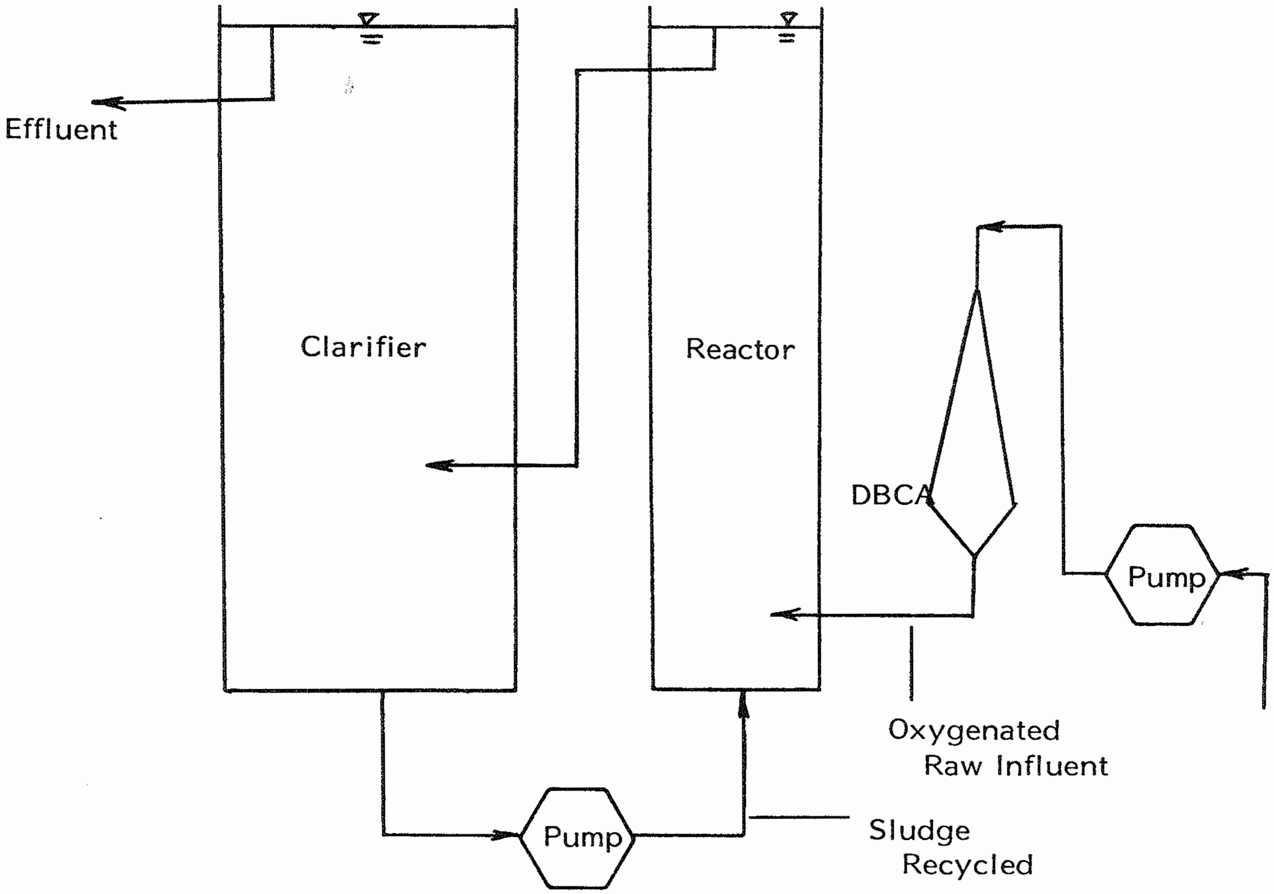

FIGURE 13. DBCA ACTIVATED SLUDGE SYSTEM, UNIT III

test meter.

ANALYTICAL METHODS AND TECHNIQUES

Suspended and Volatile Suspended Solids

Well-mixed samples were filtered through glass fiber filters which were 5.5 cm in diameter. The procedures recommended by the Environmental Protection Agency manual for methods and analysis of water and wastewater were used (32).

Chemical Oxygen Demand

Chemical oxygen demand was determined by the *Standard Methods* dichromate reflux method using silver catalyst (23). Chloride corrections were made by the addition of mercuric sulfate. Membrane filters with an effective pore size equal to 0.45 m were used for "filtered" sample determinations.

Biological Oxygen Demand

BOD determinations were performed in accordance with Section 219, *Standard Methods* (34). The feed sample was not seeded but the other BOD samples were seeded with dilution water containing 1 ml of settled raw wastewater per liter.

Dissolved Oxygen Concentration

The dissolved oxygen was monitored by a modified Galvanic Cell. The modified meter is capable of direct readings of DO = 0 to 200 mg/l. The meter was calibrated periodically by a modification of the Winkler dissolved oxygen determination test. For DO concentrations above 30 mg/l, a dilution process was necessary. The dissolved oxygen was measured by the method in Section 218B, Standard Methods, the azide modification.

Total Organic Carbon, TOC

The TOC concentrations were determined by using a Beckman Model 915 Carbon Analyzer. The TOC was standardized daily prior to analysis.

Total Oxygen Demand, TOD

The TOD concentrations were determined by using the Ionics Inc. Total Oxygen Demand Analyzer. All samples were homogenized and the TOD apparatus was standardized prior to analysis. The instrument gives direct read out of oxygen utilized in the oxidation of carbon, nitrogen, and some sulfur compounds.

V. RESULTS AND DISCUSSION

PRELIMINARY EVALUATION OF DBCA

Initial testing was performed to check the operation of the system and the oxygen absorption efficiency of the DBCA. The minimum inlet velocity acceptable to the system was approximately 8 ft/sec. Various sizes of pipes allow the inlet velocity to remain constant while the flow rate changes. The maximum workable inlet velocity is defined as the flow condition at which the bubbles remain in the DBCA and was approximately 12 ft/sec. The minimum workable inlet velocity is that which provides sufficient inlet velocity to provide for dispersion of the bubble swarm in the DBCA.

To insure maximum utilization, the best absorption efficiency per oxygen-water ratio was determined. Oxygen transfer is partial pressure dependent; therefore, two pressures (45 and 90 psig) were evaluated. The extreme design velocity rates of 8 and 12 fps were evaluated. In addition, these tests were conducted on fresh water and settled raw municipal wastewater to monitor any differential absorption resulting from surface tension.

A summary of the data observed under fresh water conditions is presented in Figure 14. Pressure seemed to have negligible effect upon absorption efficiency. The main factors affecting the transfer efficiency were oxygen-water ratio and the design velocities. Above an oxygen-water ratio of 2 percent, the controlling factor for absorption was the oxygen-water ratio; the inlet velocity rates were insignificant. Below 2 percent oxygen-water ratio, the upper design inlet velocity followed the expected linear results. However, the lower design inlet velocity gave considerably better transfer efficiency than anticipated. At one percent oxygen-water ratio the transfer efficiency for the lower inlet velocity was 96 percent. Care in operation was necessary to prevent phase separation which occurred periodically, usually at the minimum inlet velocity.

The effect of surface tension upon the transfer was negligible with the settled untreated municipal wastewater. The only difference was that at one percent oxygen-water ratio the oxygen transfer efficiency reached 98 percent for the lower inlet velocity.

DBCA ACTIVATED SLUDGE SYSTEM, UNIT I

The initial continuous flow activated sludge system incorporating DBCA (Figures 11 and 15) was a single unit, reactor clarifier. After the preliminary work on oxygen transfer was

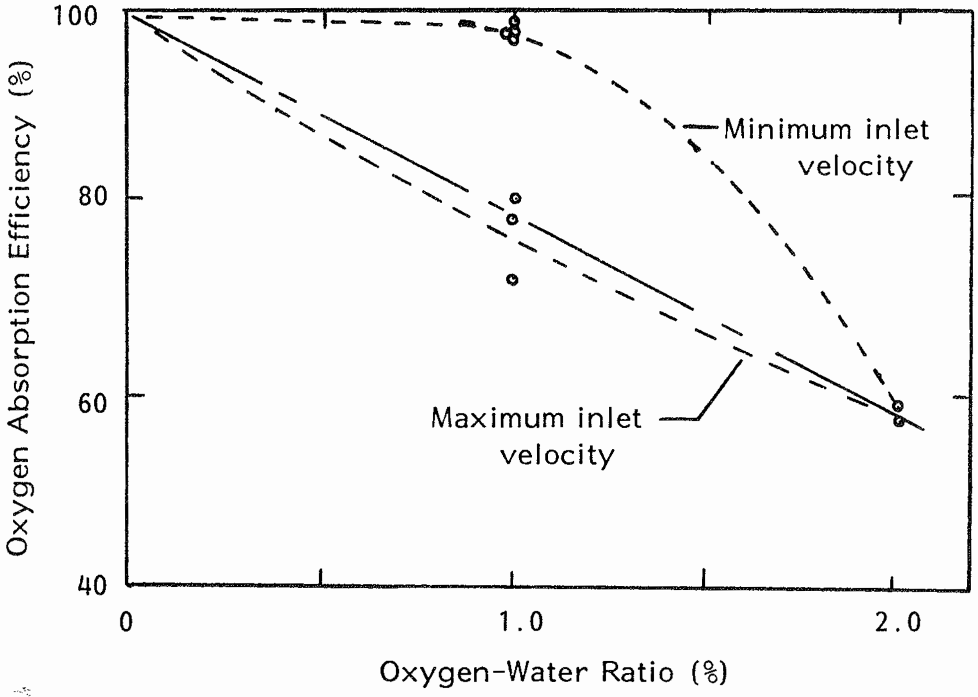

FIGURE 14. ABSORPTION EFFICIENCY OF DBCA AT VARIOUS OXYGEN-WATER RATIOS

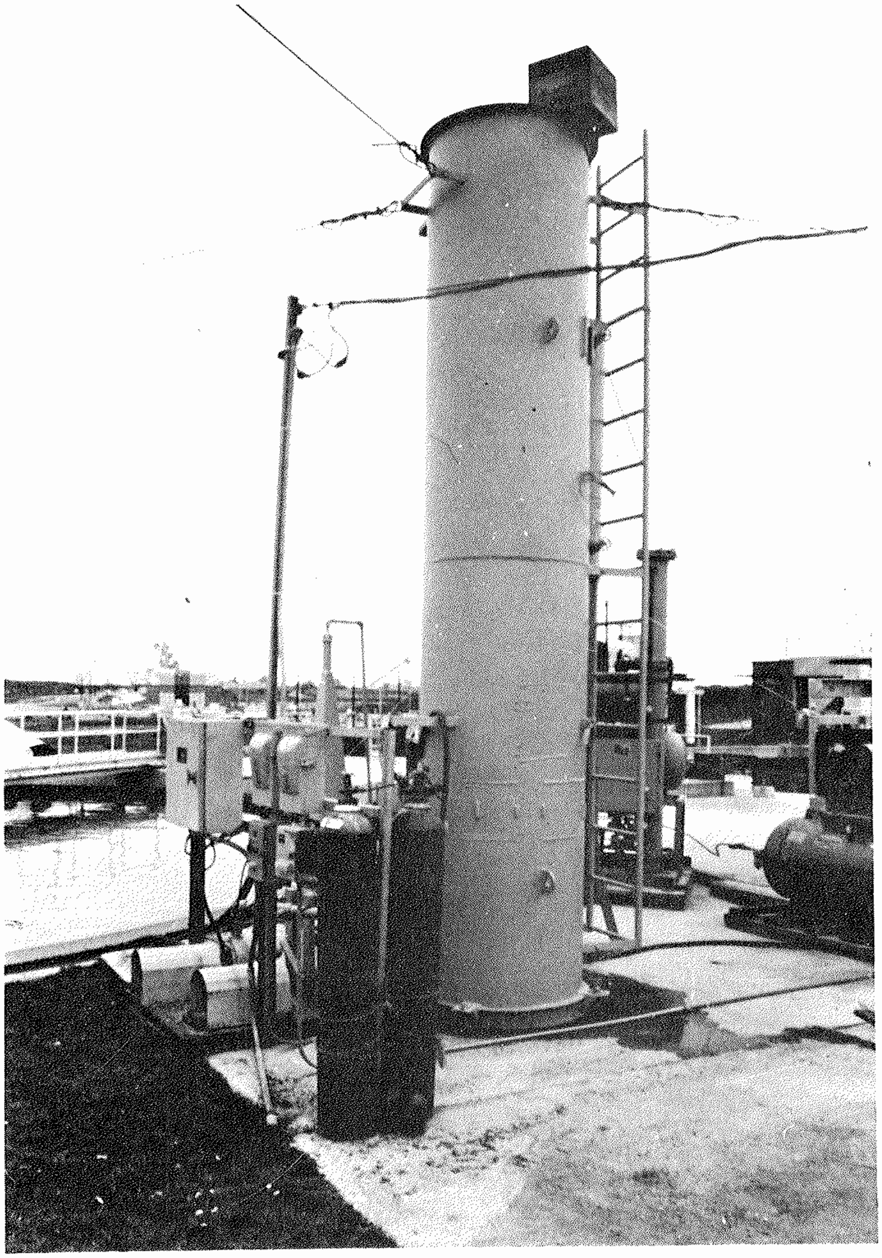

FIGURE 15. DBCA ACTIVATED SLUDGE SYSTEM, UNIT 1 (Pictoral)

completed, the unit was operated at an oxygen-water ratio of one percent which represented a transfer efficiency in excess of 95 percent. The sludge was recycled at a 5 : 1 ratio through the DBCA. The sludge recycle rate (oxygenated sludge) was a function of supplying equivalent oxygen (lb/day) to the BOD_5 (lb/day). When continuous operation was attempted, sludge flotation problems appeared immediately.

Initially, the possibility of beating up the floc and adversely affecting settling was questioned. Several settleability tests were performed and the data presented in Appendix C indicated that this was not a significant effect. As the sludge passed through the DBCA, the minute oxygen bubbles attached to the floc particles, thereby floating the material.

A physical modification of the unit (previously discussed) was employed to aid in solids and liquid separation. An artificial mat of synthetic trickling filter media was placed just above the sludge recycle line and was partially effective in physically shearing the attached bubbles from the sludge. However, the shear was not sufficient to prevent the washout of the solids from the system.

Mixing also became suspect and was intensified to increase the shear of the oxygen bubbles from the solids. Compressed air was released at various strategic locations in the chamber to produce a rolling motion and the necessary shear but the system could not continuously operate in this mode. However, this effort was futile and sludge flotation persisted.

Several attempts to use mechanical mixers were undertaken to enhance solids separation. Mixing equivalent to that observed in air activated sludge processes was applied; however, this procedure met with limited success. The flotation of the solids persisted regardless of the mixing. Therefore, the system failed because of solids washout.

DBCA ACTIVATED SLUDGE SYSTEM, UNIT II

The original reactor-clarifier was modified and the individual unit processes were separated into distinct units for more positive control (Figures 12 and 16). The system was operated with and without mechanical mixing for bubble shear.

The system was operated without auxiliary mechanical mixing, but hydraulic mixing was provided during head equalization. Although the mixing during transfer from unit to unit was extensive, some flotation appeared in the clarifier, but more as a scum. The final result again was failure because of solids washout; however, the operating time before failure was extended.

The system then was operated with hydraulic and auxiliary mechanical mixing. Sludge flotation problems were

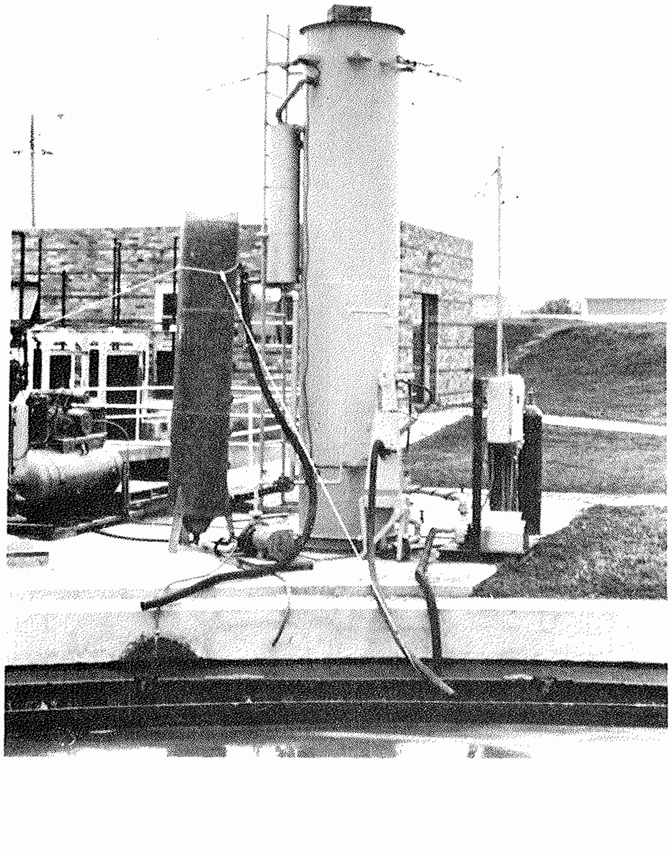

FIGURE 16. DBCA ACTIVATED SLUDGE SYSTEM, UNIT 11 (Pictoral)

essentially eliminated; however, the sludge settleability (Appendix C) was poor. The poor settling sludge resulted in a F/M ratio of 0.1 lb BOD_5/lb MLSS which also was impossible to maintain.

The sludge flotation problems were overcome but the system failed again because of solids washout or bulking.

DBCA ACTIVATED SLUDGE SYSTEM, UNIT III

This unit entailed a complete change in design. The influent was oxygenated instead of the recycled sludge mass (Figures 12 and 17). The unit was operational and no solids flotation or bulking was observed during the test period. The test period was reduced to three weeks by mechanical failure of sludge recycle pumps, rather than solids washout.

Operation and Problem Areas

The DBCA system was particularly difficult to operate continuously for extended periods of time (20 days). Most of the problems were due to minor fluctuations in flow rates or pressures. At times the pressure rose above the inlet oxygen pressure, causing back flow, and resulting in short shut downs of one hour or less which were required to clean and clear the oxygen flow meter. This problem occurred only while operation personnel were on site.

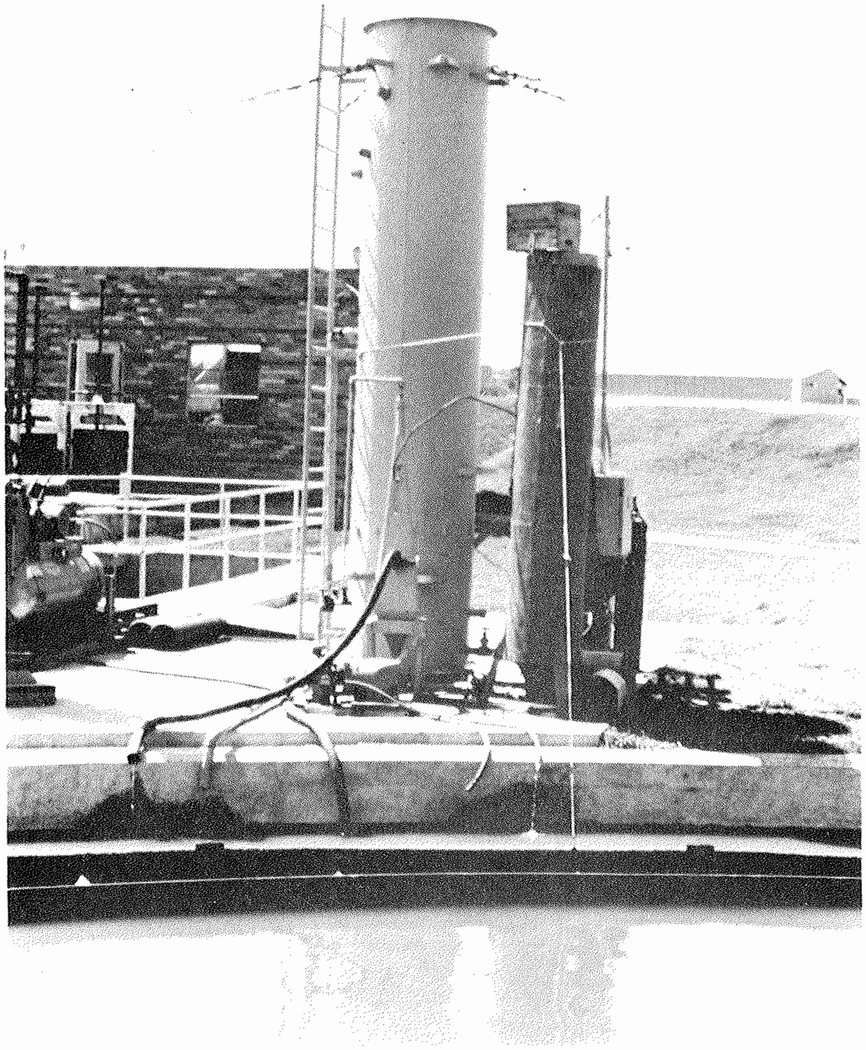

FIGURE 17. DBCA ACTIVATED SLUDGE SYSTEM, UNIT 111 (Pictoral)

Phase separation would sometime occur after continuous unattended operation. This problem caused some concern because the operation was often unattended for 12 to 24 hours. Therefore, the sampling schedule included collection of samples over a 6 to 8 hour period followed by adjustment of the instruments for the nighttime operation. The system operated some nights without problem; however, occasionally (twice a week) during the unattended period, phase separation would occur. But this problem was not considered serious other than pointing out the need for automation and/or continual supervision.

A troublesome area, which was actually more difficult to overcome than operation problems with DBCA, was sludge recycle. Sludge recycle pumps failed periodically throughout the study and finally resulted in the premature conclusion of the study. However, a problem which resulted in difficult data evaluation was the inability to collect and recycle settled sludge from the clarifier. The vessel used as a clarifier in this portion of the study was the reactor in Unit II and had no sludge collection device. Thus, any heavy settled matter near the tank periphery was lost and considered wasted. Occasional rapid sludge recycle to reclaim some of this material and some additions of fresh activated sludge were employed. A discussion of the mass balance necessary for analysis is presented in the evaluation of kinetics.

Many unforeseen and difficult operational problems occurred. Most resulted from the device (DBCA) being unattended during most of the operations. However, the initial design, which resulted in sludge flotation, was a burden to the successful completion of the project. Problems, such as sludge collection which was unnecessary in Unit I, occurred because equipment initially selected to satisfy requirements was not readily adaptable to changes in design for Unit III.

Growth and Substrate Removal Kinetics

Growth kinetics could be modeled after the simplified mass balance presented in Equation 5 since the same soluble substrate was used throughout the study.

$$D_c = 1/\theta_c = Yq - k_d \quad . \quad . \quad . \quad . \quad . \quad . \quad . \quad . \qquad 5$$

where:

D_c is the net specific growth rate or cell dilution rate in lb excess VSS produced per day per lb MLVSS.

Y is the yield coefficient in lb VSS produced per lb BOD removed.

q is the substrate removal rate in lb BOD_5 removed per day per lb MLVSS.

k_d is the specific endogenous respiration coefficient in day .

θ_c is the mean solids retention time in days.

The values presented in Appendix E of every fourth day's data were plotted on a graph of $1/\theta_c$ versus the substrate removal rate q. A straight line of best fit was constructed by least squares through the data and the yield coefficient, Y, determined from the slope of the line. The intercept determined the decay rate, k_d . The growth plots were constructed on both BOD_5 and COD bases and are presented in Figure 18. The cyclic changes in concentration of the raw wastewater allowed the loadings to change sufficiently for the curve to be constructed over a narrow range.

A mass balance on solids was necessary to calculate volatile suspended solids wasted because of the uncontrolled wastage of the sludge. An example of this calculation is presented in Appendix E. The only meaningful numbers are the ratio of VSS wasted and total MLVSS in the system.

The yield coefficient was 0.46 pound volatile suspended solids produced per pound COD removed with a decay rate of 0.02 per day. On a BOD_5 basis the yield was 0.73 pound volatile suspended solids produced per pound BOD_5 removed with a decay rate of 0.06 per day. The high yield on the BOD_5 basis may be attributed to the rather low influent BOD_5 concentration.

Equation 4 was used to calculate the substrate removal kinetics. The reciprocal of the terms are presented in Equation 7. A graphical presentation of Equation 7 results in a

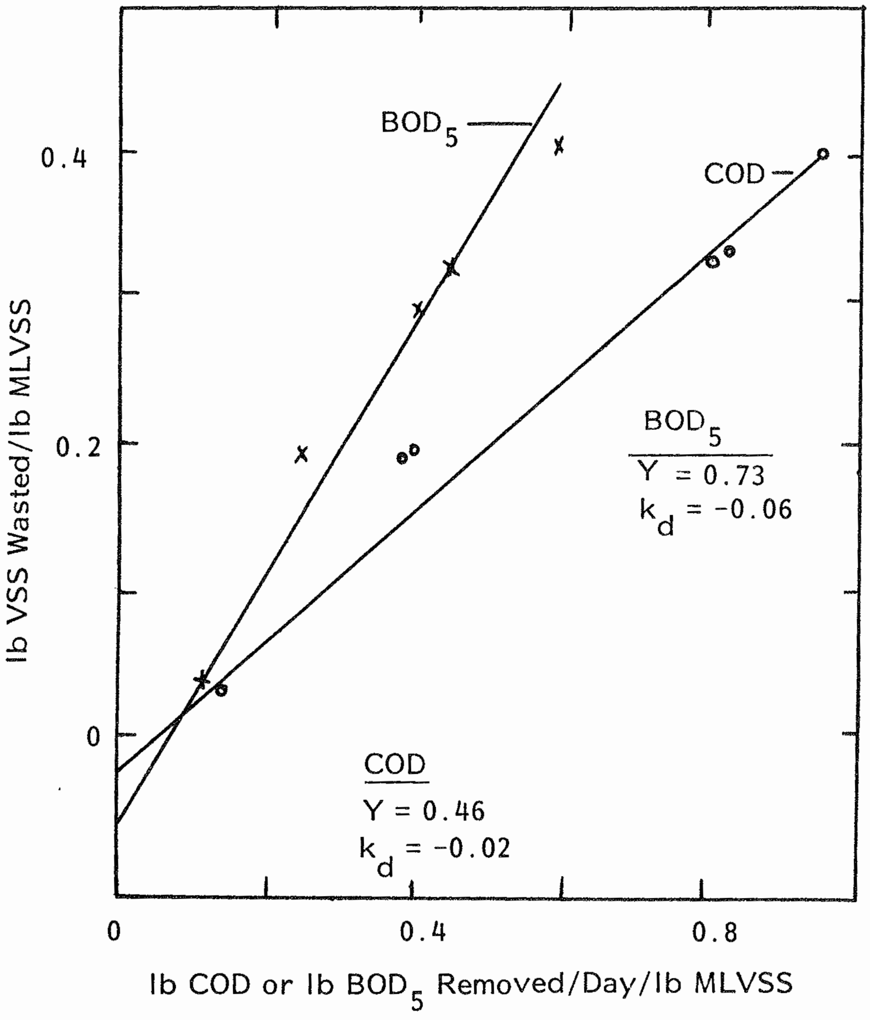

FIGURE 18. GROWTH KINETICS

straight line relationship.

$$q = \mu/Y \; S/K_s + S \quad . \; . \; . \; . \; . \; . \; . \quad 4$$

$$1/q = K_s/\mu/Y \; (1/S) + 1/\mu/Y \quad . \; . \; . \; . \; . \quad 7$$

where:

q	is the rate of substrate removal in lb BOD removed per day per lb MLVSS.
K_s	is the substrate concentration mg/l at half the maximum growth rate.
μ	is the net specific growth rate in lb VSS produced per lb VSS.
Y	is the yield coefficient in lb VSS produced per lb BOD removed.
S	is the substrate concentration in mg/l.

The relationship for the data observed in this study is presented in Figure 19 and complete data are included in Appendixes D and E. A least squares computation was used to obtain the line of best fit. The maximum net specific growth rate, , was 3.7 lb VSS produced per lb VSS based on COD, and the substrate concentration at half the maximum growth rate was 281 mg/l of COD. A summary of kinetic coefficients and growth rates is presented in Table IV.

All kinetic data agreed well with published data presented in Table V. The sludge production compares well with data reported for other high purity oxygen plants which are presented in Table VI. It should be noted that DBCA system

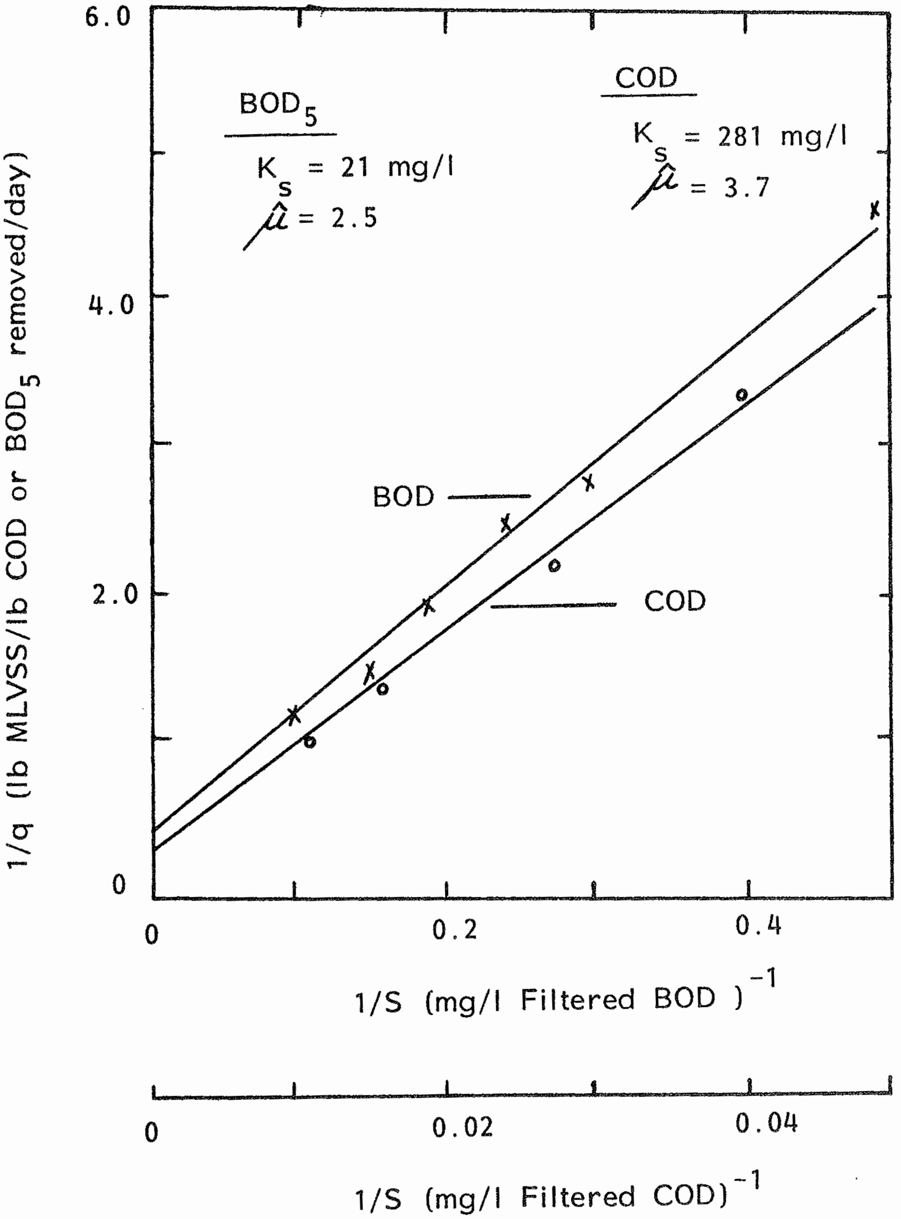

FIGURE 19. SUBSTRATE REMOVAL KINETICS

TABLE IV.

SUMMARY OF KINETIC COEFFICIENTS*

Coefficient	Basis	
	COD	BOD_5
Y, lb VSS produced/lb substrate	0.46	0.73
K_d, days	0.02	0.06
$\hat{\mu}$, lb VSS produced/lb VSS	3.7	2.5
K_s, mg/l	281	21

* At Temperature = 20 plus or minus 2 degrees
Centigrade and pH = 7.3

TABLE V.

SUMMARY OF REPORTED KINETIC DATA FOR MIXED CULTURE AEROBIC SYSTEMS

Substrate	Base	Y	k_d	μ	K_s	Temp ^oC	Investigator	Ref.
Glucose	BOD	.42	.09	1.3	355	-	Stack & Conway	35
	BOD	.42	.008	1.3	355	-	Pearson	36
	BOD	.59	-	1.9	-	10	Garrett & Sawyer	37
Glucose-peptone	BOD	.49	-	5.1	-	20	Garrett & Sawyer	37
Peptone	BOD	.43	-	6.2	65	30	Garrett & Sawyer	37
Skim Milk	BOD	.48	.045	2.4	110	-	Gram	38
	COD	.43	.061	2.9	110	-	Eckhoff & Jenkins	39
Synthetic Waste	BOD	.65	.18	-	-	-	McCarty & Brodersen	40
	COD	.46	.17	-	-	-	Eckhoff & Jenkins	39
	COD	.44	.12	2.0	312	24	Ball, J.E.	16
Domestic Sewage	BOD	.50	.055	-	-	-	Heukelekian	41
	BOD	.67	.048	-	-	20	Middlebrooks & Garland	42
	BOD	.60	.019	-	-	-	Pearson & Haas	43
	BOD	-	.038	-	-	-	Garland & Shell	44
	COD	.67	.07	3.8	22	-	Benedek & Horvath	45
	COD	.35	.012	-	-	-	Pearson & Haas	43
	COD	.33	.75	-	-	-	Eckhoff & Jenkins	39
	COD	-	.055	-	-	-	Garland & Shell	44
	COD	.46	.172	-	-	-	Eckhoff & Jenkins	39
	COD	.78	-	-	-	-	Jenkins & Menar	46
	BOD	1.00	.40	12.5	180	-	Ball, R.O.	17

TABLE VI

PREVIOUSLY REPORTED SLUDGE YIELD COEFFICIENTS

Location	Y, lb VSS produced/lb BOD_5 removed
Batavia	0.4 - 0.6
Newtown	0.51
Brewery	0.56
Blue Plains	0.38 - 0.47
Simplox	0.23
Other UNOX	0.41,0.48

influent BOD_5 and COD were considerably lower than others reported which may explain the slightly higher BOD_5 yield.

Sludge Indexes

A detailed analysis of sludge settleability was beyond the scope of this study. All sludge data were related to the SVI, which was employed as a comparison unit with previous settling data in the literature although SVI is not very well suited to design. The average SVI for the 20 days of operation was 84 ml/g at 2872 mg/l MLVSS and a food to microorganism ratio of 0.42 lb BOD_5/lb MLVSS.

The average solids retention time, SRT, was 4.5 days at an overflow rate of 410 gpd/sq ft and a hydraulic detention time of 1 hour.

These data are generally in agreement with other data for high purity oxygen activated sludge systems (Table VII). The SVI reported for the DBCA of 84 ml/g is slightly greater than the other units operated at corresponding F/M ratios; however, the SVI generally is a function of MLSS and SRT which could explain the lower values reported at Newtown and in the Brewery studies.

The sludges developed with the DBCA and SIMPLOX systems compared favorably. The overflow rates, SVI, and MLSS

TABLE VII

REPORTED SLUDGE CHARACTERISTICS AND LOADINGS

Location	F/M lb BOD_5/lb	Overflow Rate gpd/sq ft	SVI ml/g	MLVSS mg/l	SRT days
Newtown	0.67	977	42	3960	3.1
Brewery	0.56	-	50	-	-
Blue Plains	0.50	-	-	-	5.0
Simplox	0.07	417	99	2950	78.0
DBCA	0.42	410	84	2872	4.5

were essentially the same. The F/M ratio in the SIMPLOX unit was small; the weak waste used in the DBCA study may account for the difference in loadings. The only unexplainable difference between the two systems is the solids retention time, with SIMPLOX operating for a very high and unusual 78 days.

Oxygen Transfer

The oxygen transfer efficiency of DBCA was maintained at 95 to 98 percent. The transfer efficiency was checked periodically throughout the study period. No problems were encountered in maintaining a 95 percent transfer efficiency at 1 percent oxygen-water ratio by volume.

The oxygen transfer of DBCA compares favorably with the covered tank operations at Batavia, Newtown, and Blue Plains. In these studies, the computed oxygen transfers were 92, 96, and 97 percent, respectively.

Dissolved Oxygen Levels Maintained in System

The dissolved oxygen levels at selected points in the system were monitored periodically throughout the study. The influent wastewater should have a dissolved oxygen near zero; an average DO of 0.2 mg/l was recorded. An average of 85 mg/l was obtained after passage through the DBCA. It was assumed that all immediate oxygen demand was satisfied and that the DO transferred was sufficient for soluble BOD_5 reduction. This assumption proved valid since the average DO of the effluent from the reactor was 7.0 mg/l. The final effluent at the point of discharge contained an average DO concentration of 4.5 mg/l.

Other Pertinent Information

The air temperature ranged from 5 degrees Centigrade to 33 degrees Centigrade during the operation of the system; the water temperature remained fairly constant at 20 degrees Centigrade plus or minus 2 degrees Centigrade. The municipal wastewater had a high buffering capacity and the pH throughout the operation remained around 7.3.

Removal efficiencies of COD, BOD_5, and suspended solids are presented in Table VIII. These reductions compare reasonably well with other reported data presented in the literature review. The suspended solids removal was low because the influent to the system was settled municipal wastewater containing very light solids. Also, some of the removals were affected by the low soluble BOD_5 of the food supply.

TABLE VIII

DBCA REMOVAL EFFICIENCIES

Parameter	Percent Removal
Average Total COD	80.4
Average Soluble COD	64.1
Average Total BOD_5	94.5
Average Soluble BOD_5	90.3
Average Suspended Solids	70.0

VI. CONCLUSIONS

1. The incorporation of Downflow Bubble Contact Aeration into an activated sludge system is feasible and an operational system was achieved with treatment performance equal to other larger scale systems reported in the literature.

2. Kinetic coefficients were determined; they compared favorably with previously reported values. These parameters are: (a) sludge yield, Y = 0.47 lb volatile suspended solids per lb COD removed, (b) endogenous respiration coefficient k_d = 0.02 per day, (c) maximum net specific growth rate, μ = 3.7 lb VSS produced per lb VSS, and (d) the substrate concentration at half the maximum net specific growth rate, K_s = 281 mg/l.

3. The BOD_5 and COD removed by the DBCA system was 94.5 and 80.4 percent, respectively. The system removals were as good as any previously reported values. Solids separation was a recurring problem, as in most activated sludge plants, during initial testing because of sludge flotation. After modifications to the system eliminated flotation, solids separation appeared to be excellent.

4. The design of the DBCA needs to be changed to

incorporate automatic monitoring and control equipment to insure longer unattended, uninterrupted service. Design of DBCA units may be somewhat limiting because of the inflexibility to severe changes in the loadings. This variability in performance might be overcome by the addition of an equalization basin to eliminate high peak loadings. One immediate design change might include some kind of a pop-off valve to reflood the chamber should phase separation occur. Care in design should be utilized in an attempt to eliminate phase separation. Further research may also prove to be beneficial.

5. Better initial pilot plant design would improve the operation and possibly improve the efficiency of treatment. Also some flexibility should be incorporated to enable a greater depth of study into solids settling, solids retention time, and higher loadings. Corrosion of DBCA also should be anticipated and monitored.

6. A larger scale DBCA activated sludge unit, especially one which is fully automated, should be investigated. The practical application of DBCA seems to await a debugging of the process and a greater scope of research.

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APPENDIX

D.B.C.A. EVALUATION

Type Water	Oxygen-Water Ratio (percent)	Pressure (psig)	Inlet Velocity (fps)	Dissolved Oxygen @ 100% Absorption (mg/l)	Actual Dissolved Oxygen (mg/l)	Transfer Efficiency (percent)
Fresh	1	45	12	53	41	77
Fresh	1	90	12	92	65	70
Fresh	1	45	8	53	51	96
Fresh	1	90	8	93	89	96
Fresh	2	45	12	106	57	54
Fresh	2	90	12	186	106	57
Fresh	2	45	8	106	59	56
Fresh	2	90	8	186	97	52
Sewage	1	45	8	53	52	98
Sewage	1	45	12	53	43	79
Sewage	1	90	8	93	90	97
Sewage	2	90	12	186	106	57

EXAMPLE CALCULATIONS DETERMINING THEORETICAL DISSOLVED OXYGEN AND FLOW RATE

Water flow rate for inlet velocity of 12 fps = 17 gpm
= .0378 cfs

At 1 percent Oxygen-Water Ratio, Respective Pressure, and Temperature

@ 0 degrees C
15 psia
0 psig

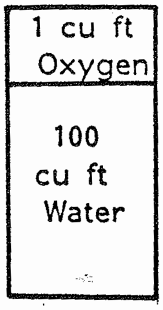
$$\frac{359 \text{ cu ft Oxygen}}{32 \text{ lb}} = \frac{11.2 \text{ cu ft Oxygen}}{\text{lb}}$$

@ 20 degrees
15 psia
0 psig

$$\frac{11.2(273 - 20)}{273} = \frac{12 \text{ cu ft Oxygen}}{\text{lb}}$$

@ 20 degrees C
60 psia
45 psig

$$\frac{15(12)}{60} = \frac{3 \text{ cu ft Oxygen}}{\text{lb}}$$



DO for 100 percent absorption at
1 percent Oxygen-Water Ratio

$$100(7{,}500{,}000)\text{DO}(8.34) = \frac{1 \text{ cu ft}}{3 \text{ cu ft/lb}}$$

$$\text{DO} = 53 \text{ mg/l}$$

2 percent Oxygen-Water Ratio

$$100(7{,}500{,}000)\text{DO}(8.34) = \frac{2 \text{ cu ft}}{3 \text{ cu ft/lb}}$$

$$\text{DO} = 106 \text{ mg/l}$$

Oxygen Flow Rate at 1 percent Oxygen-Water Ratio and 45 psig

Water Flow = 0.378 cfs
Oxygen Flow = .000378 cfs

$$\text{Oxygen Flow @ STP} = \frac{60(.000378 \text{ cu ft})(28.3 \text{ l})(60 \text{ sec})}{15 \text{ sec-cu ft-min}}$$

$$= 2.56 \text{ l/min}$$

SLUDGE SETTLEABILITY TESTS*

Time (min)	Raw Seed Sludge Height (ml)	Sludge Passed Through High Press. Height (ml)	Sludge Passed Through DBCA At Low Oxygen Flow Height (ml)	Unit II Activated Sludge Height (ml)
0	1000	1000	1000	1000
5	800	925		1000
10	600	750	Air Floated	1000
15	450	550	To Top	950
20	300	470		900
30	175	250		750

* Conducted in 1000 ml graduated cylinder

DATA SUMMARY

	Total Organic Carbon			Total Oxygen Demand		
Date	Infl. (mg/l)	Filtr. Infl. (mg/l)	Filtr. Effl. (mg/l)	Infl. (mg/l)	Filtr. Infl. (mg/l)	Filtr. Effl. (mg/l)
3 - 17	80	72	10	125	71	50
3 - 18	78	68	10	130	77	50
3 - 19	76	71	10	148	110	50
3 - 20	83	72	10	154	116	50
3 - 21	90	84	49	192	139	106
3 - 22	100	96	67	218	186	50
3 - 23	86	76	10	132	128	50
3 - 24	88	86	10	161	128	50
3 - 25	104	97	76	252	195	89
3 - 26	97	86	76	252	182	99
3 - 27	90	88	33	208	178	50
3 - 28	86	78	24	192	161	50
3 - 29	107	85	52	258	161	50
3 - 30	73	62	35	182	171	50
3 - 31	65	55	32	148	127	50
4 - 1	100	85	52	241	208	101
4 - 2	97	87	42	259	243	109
4 - 3	104	101	24	267	249	75
4 - 4	100	92	30	252	237	68
4 - 5	106	89	24	238	220	63

DATA SUMMARY

	Chemical Oxygen Demand			Biological Oxygen Demand		
Date	Infl. (mg/l)	Filtr. Infl. (mg/l)	Filtr. Effl. (mg/l)	Infl. (mg/l)	Filtr. Infl. (mg/l)	Filtr. Effl. (mg/l)
3 - 17	58	43	19	57	27	2
3 - 20	105	74	27	73	44	5
3 - 23	148	82	16	79	48	3
3 - 26	217	148	62	119	88	10
3 - 29	188	72	32	92	43	4
4 - 1	308	136	44	126	60	4
4 - 4	241	153	57	103	74	6

DATA SUMMARY

Date	Influent		Effluent		Mixed Liquor		Recycle Solids	
	SS (mg/l)	VSS (mg/l)	SS (mg/l)	VSS (mg/l)	SS mg/l)	VSS (mg/l)	SS (mg/l)	VSS (mg/l)
3 - 17	50	45	15	15	4550	3550	9000	6875
3 - 18	105	95	35	30	3580	2800	8425	6380
3 - 19	65	50	25	25	4080	3120	7950	5710
3 - 20	92	75	20	20	4200	3060	7325	5500
3 - 21	100	95	30	30	3780	2950	7040	5115
3 - 22	110	100	35	30	3570	2820	6610	4900
3 - 23	35	30	20	15	3510	2730	5950	4700
3 - 24	50	40	15	15	3270	2585	6365	4700
3 - 25	95	90	30	30	3270	2500	6020	4625
3 - 26	75	70	25	25	3725	2930	6150	4940
3 - 27	95	85	25	20	3650	2730	6080	4785
3 - 28	85	75	35	35	3700	2750	5975	4565
3 - 29	80	70	30	25	3450	2695	6295	4860
3 - 30	60	55	20	20	3500	2685	6015	4725
3 - 31	45	35	20	20	3325	2700	5885	4695
4 - 1	130	120	20	20	4160	3160	7090	5300
4 - 2	95	75	40	35	4000	2990	6750	5175
4 - 3	100	85	30	25	3850	2950	6700	4935
4 - 4	125	100	25	25	3790	2815	6550	4500
4 - 5	85	80	35	30	3610	2775	6310	4375

SAMPLE CALCULATIONS FOR LOADINGS AND MASS BALANCE DETERMINATION FOR SOLIDS WASTING

Loading

$$\text{Volume of sludge} = \frac{(1.5)\,(1.5)\,(8)\,(7.48)}{4} = 105.7 \text{ gal.}$$

$$\text{Avg. lb MLSS} = \frac{(105.7)\,(3728)\,(8.34)}{1{,}000{,}000} = 3.29 \text{ lb}$$

$$\text{Avg. lb MLVSS} = \frac{(105.7)\,(2872)\,(8.34)}{1{,}000{,}000} = 2.53 \text{ lb}$$

$$\text{Avg. lb BOD removed} = \frac{2\,(1440)\,(99-5)\,(8.34)}{1{,}000{,}000} = 2.25 \text{ lb}$$

$$\text{Avg. lb soluble BOD removed} = \frac{2\,(1440)\,(64-5)\,(8.34)}{1{,}000{,}000} = 1.41 \text{ lb}$$

$$\text{Avg. lb COD removed} = \frac{2\,(1440)\,(211-43)\,(8.34)}{1{,}000{,}000} = 4.03 \text{ lb}$$

$$\text{Avg. lb soluble COD removed} = \frac{2\,(1440)\,(118-43)\,(8.34)}{1{,}000{,}000} = 1.80 \text{ lb}$$

Avg. F/M = 0.55 lb soluble BOD_5/lb MLVSS

Avg. F/M = 0.71 lb soluble COD/lb MLVSS

Mass Balance

All relative so that only the ratio, lb wasted/ lb MLVSS, is an actual number.

Amount wasted = MLVSS - Recycle

$$= \frac{4(3550) - 2(6875)}{2} = 225$$

lb wasted/lb MLVSS = 225/3550 = 0.06

LOADINGS AND SOLIDS WASTING

Date	lb MLVSS	lb BOD removed	lb COD removed	$\frac{\text{lb COD removed}}{\text{lb MLVSS}}$	$\frac{\text{lb BOD removed}}{\text{lb MLVSS}}$	$\frac{\text{lb wasted}}{\text{lb MLVSS}}$
3 - 17	3.13	0.58	0.58	0.19	0.18	0.06
3 - 20	2.70	0.94	1.13	0.42	0.35	0.20
3 - 23	2.41	1.08	1.59	0.66	0.45	0.28
3 - 26	2.58	1.87	2.06	0.80	0.73	0.32
3 - 29	2.38	0.94	0.96	0.40	0.39	0.19
4 - 1	2.79	1.34	2.21	0.79	0.48	0.32
4 - 4	2.48	1.63	2.30	0.93	0.66	0.40