# EPA RESEARCH AND DEVELOPMENT ACTIVITIES WITH OXYGEN AERATION

## TECHNOLOGY TRANSFER DESIGN SEMINAR

U.S. ENVIRONMENTAL PROTECTION AGENCY
OFFICE OF RESEARCH AND MONITORING
NATIONAL ENVIRONMENTAL RESEARCH CENTER
CINCINNATI, OHIO

# EPA RESEARCH & DEVELOPMENT ACTIVITIES WITH OXYGEN AERATION

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Office of Research & Monitoring
Cincinnati, Ohio

# EPA Research & Development

# Activities with Oxygen Aeration

by

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## INTRODUCTION

The concept of using oxygen enriched air or pure oxygen as the aeration gas in the activated sludge process dates back over twenty years to Pirnie and Okun's "bio-precipitation" model. Since then, many investigators have acknowledged that oxygen gas has inherent characteristics which could be used to advantage in activated sludge processing. On the other hand, during this same period, oxygen aeration was not considered economically practical because existing gas contacting mechanisms were not capable of effectively dissolving and utilizing the relatively expensive oxygen commodity.

In 1968, a $528,000 research contract was awarded by EPA's forerunner, FWPCA, to the Linde Division of the Union Carbide Corporation to evaluate an oxygen aeration system which offered promise for overcoming the poor utilization factor. This project (herein after referred to as Batavia I) was carried out in 1969 at the Batavia, New York Municipal Pollution Control Plant (nominal capacity - 2.5 mgd) and served as the foundation which stimulated the exceptional growth and development of this process. In the short span of three years, oxygen aeration has come full cycle from a research undertaking to the threshold of widespread acceptance and utilization by the municipal and industrial waste treatment fields.

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In addition to Batavia I, EPA has sponsored four other projects involving various aspects of oxygen aeration. They include the following:

1. A follow-on research contract to Union Carbide for $209,000 primarily to investigate handling of oxygen sludges (carried out at Batavia in 1970 and hereinafter referred to as Batavia II).
2. A $1.5 million demonstration grant to New York City to demonstrate and evaluate the Union Carbide oxygen aeration system on a 20 mgd scale at the Newtown Creek Treatment Plant.
3. A $160,000 R&D grant to the City of Las Virgenes, California to evaluate a different oxygen dissolution concept developed by Cordon International (formerly Cosmodyne Corporation).
4. An in-house research project conducted at the EPA/DC Pilot Plant in Washington, D.C. for the past two years (hereinafter referred to as the Blue Plains Project).

The results of the two Batavia projects are discussed briefly in the following pages. Union Carbide will also refer to these projects in their presentation. A description of the Newtown Creek demonstration project which is just now in the final equipment checkout and system start-up phases will be given prior to the plant tour on Thursday, March 2. The Las Virgenes project is also just beginning system checkout and, not yet having generated any operating data, will only be explained briefly at this time. Major emphasis will be devoted in this text to a general discussion of the on-going 100,000 gpd Blue Plains project. This is the longest continuous oxygen aeration study undertaken at one site to examine design parameters of the process. The Blue Plains data accumulated and refined over the two-year test period represents

the major contribution which the Agency can add to this design seminar.

GENERAL DISCUSSION - BATAVIA I and II

A flow diagram of the Batavia, New York Municipal Pollution Control Plant is shown in Figure 1. The plant consists of two mirror-image treatment trains (nominal capacity - 1.25 mgd each) with separate secondary clarification and sludge recycle facilities. The Batavia Plant does not utilize primary settling of wastewater. A six-stage oxygen aeration system was installed in one of the two aeration tanks. Liquid oxygen was stored on site and vaporized prior to introduction to the first stage. The performance of the oxygen aeration system was evaluated and compared to that of the parallel conventional plug flow diffused air aeration system over a seven month period during Batavia I.

The oxygen aeration system which was installed at Batavia is illustrated schematically in Figure 2. A gas tight cover is utilized to prevent venting to the atmosphere. The aeration tank is segmented into stages by vertical baffle walls. Each stage is equipped with a recirculating compressor and a combination submerged turbine-rotating sparger. Oxygen gas is fed to the first stage and along with gaseous decomposition products and inert gases recirculated in each succeeding stage. Oxygen feed rates to the first stage and gas exhaust rates from the last stage are automatically controlled.

The pertinent results of the Batavia I project are summarized below:

1. The oxygen aeration system exhibited excellent oxygen aeration transfer capabilities with overall utilization efficiencies in excess of 90%.
2. The power required for oxygen gas dissolution was 1/5 to 1/6 of that required by the air blowers for the parallel air system.

3. The oxygen system was able to sustain high mixed liquor suspended solids (MLSS) concentrations in the order of 6000-7000 mg/l in a highly aerobic environment (8-10 mg/l mixed liquor dissolved oxygen) with relative ease. The highest MLSS concentration that could be maintained by the coarse bubble diffused air system without causing anoxic conditions was about 3600 mg/l.
4. The high solids carrying capabilities of the oxygen system permitted effective treatment of the incoming raw wastewater (average BOD, 160-260 mg/l) with nominal detention times of 1 1/2 hours (based on raw flow) and volumetric organic loadings greater than 200 lb BOD /day/ 1000 $ft^3$ mixed liquor, conditions which would constitute an excessive overload to typical air system reactors.
5. Equivalent treatment was provided by the oxygen system to that of the air system in 1/3 as much aerator volume.
6. The oxygen system produced only 50-60% as much excess biological sludge as the air system under the conditions tested.
7. Both systems exhibited good sludge settling and compaction characteristics. Average daily clarifiers loadings up to 1500 gpd/$ft^2$ could be sustained without effluent deterioration. This is at least partially attributed to the absence of primary clarification and the resulting denser sludge. Neither Batavia project was conducted during the winter when sludge settling characteristics for any biological system can be expected to be their poorest.

Oxygen sludge handling characteristics were evaluated on a pilot-scale in the Batavia II project. The most significant result of this follow-on project

WATER POLLUTION CONTROL PLANT, CITY OF BATAVIA, N.Y.
SCHEMATIC FLOW DIAGRAM

# FIGURE 1

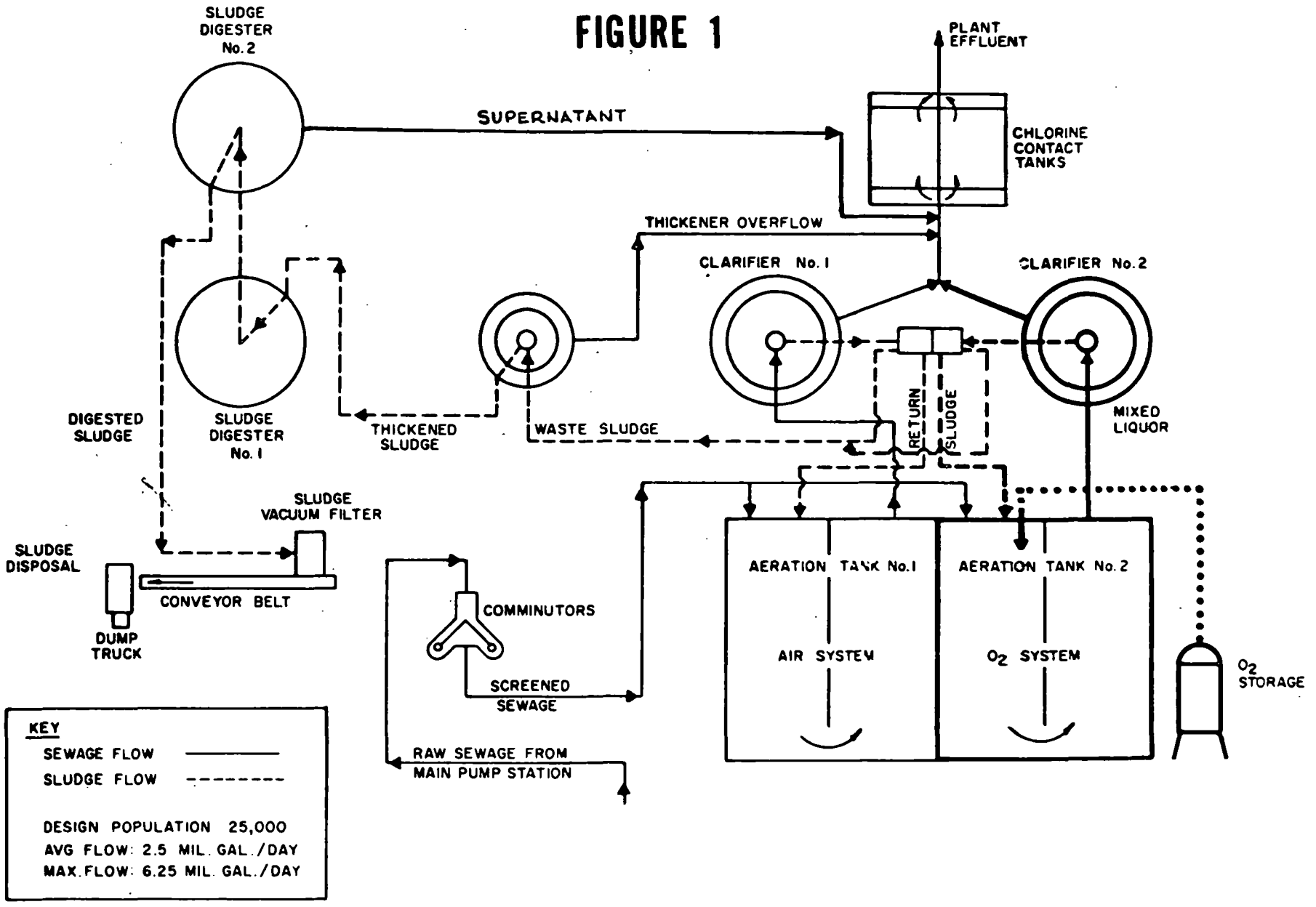
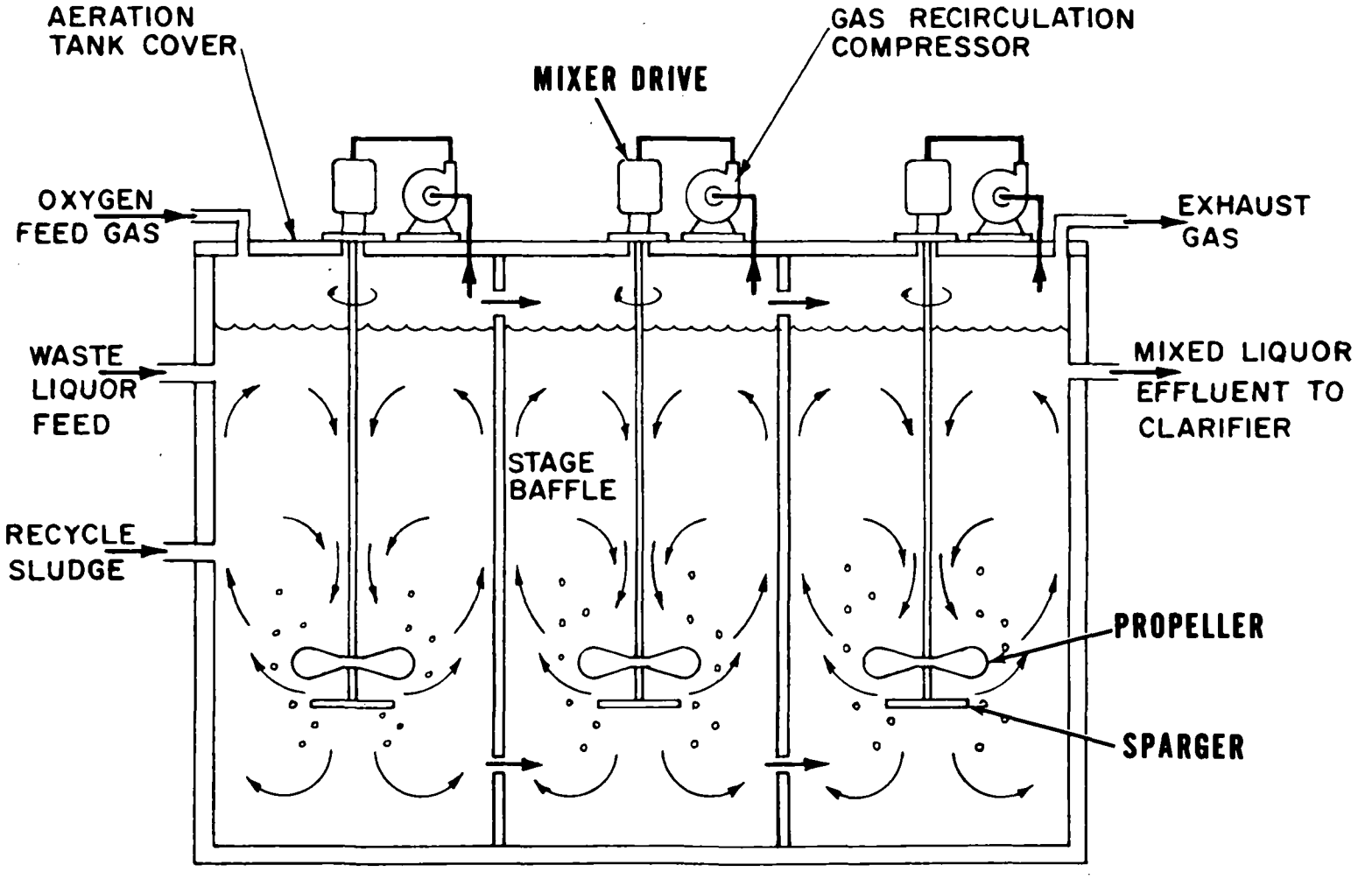


FIGURE 2

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGENATION SYSTEM UTILIZING TURBINE-SPARGERS & RECIRCULATING COMPRESSORS



was discovering that settled oxygen mixed liquor withdrawn from the clarifier underflow could be satisfactorily dewatered on a vacuum filter without prior thickening or digestion. Cake yields up to 4.5 $lb/hr/ft^2$ were obtained on a 2% solids feed at a cycle time of about 2.5 minutes/revolution. The best chemical conditioner was ferric chloride at a dosage rate of approximately 200 lb/ton of feed solids. Under these conditions, the solids content of the cake averaged around 15%. Data obtained through the courtesy of the City of Milwaukee indicated that these cake yields are 40-45% higher than Milwaukee achieves on its gravity thickened air generated waste activated sludge, with the same filter parameters and chemical conditioner.

Data from the two Batavia projects were used to develop estimated total treatment costs for an oxygen aeration system and a conventional diffused air aeration system. The costs shown in Figure 3 in cents/thousand gallons treated were estimated for new plants and include operation, maintenance and capital amortization (5-1/2%, 25 yrs.) costs for primary treatment, secondary treatment, and sludge handling and disposal facilities. In the 1 mgd range, there was projected to be no cost advantage with either system. Based on the given design assumptions, as plant size increases, cost savings begin to accrue to the oxygen system and are projected to amount to 15-20% at the 100 mgd plant size.

## GENERAL DISCUSSION - LAS VIRGENES PROJECT

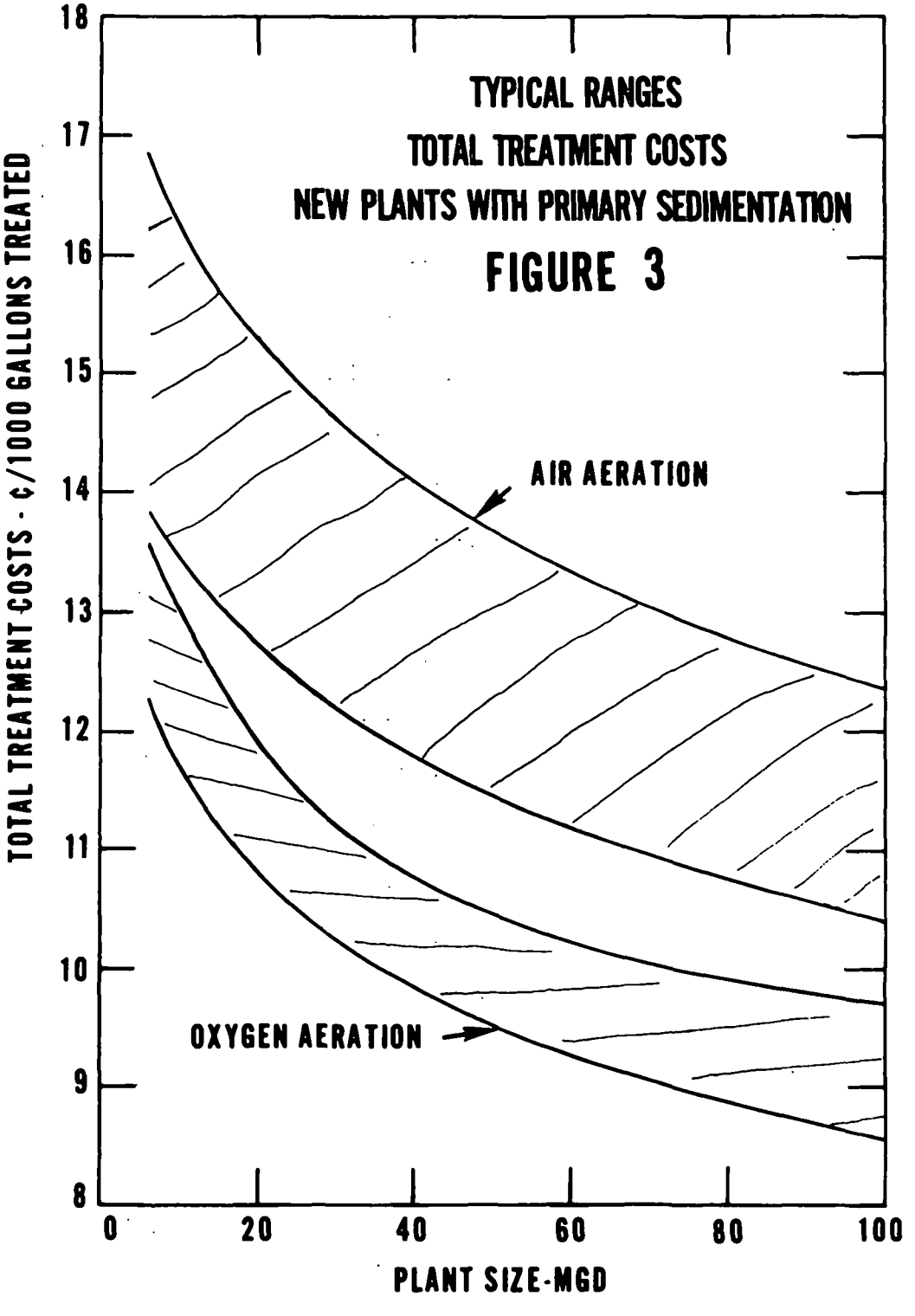
The oxygen aeration system being installed at a 2 mgd scale in the Las Virgenes, California Treatment Plant and shown schematically in Figure 4 is intended for application to existing activated sludge plants equipped with centrifugal air compressors and air diffusers. Oxygen gas is introduced through a sparger to the head of the aeration tank. Gas is continuously recycled with

the converted air aeration equipment. Exhaust gas is bled from the system at the exit end of the aeration tank.

The objective of this project is to demonstrate that oxygen aeration can be used successfully to increase organic loading capacity of existing plants with minimum capital expenditure. The additional investment consists of a tank cover (which at Las Virgenes will be an inflated dome type structure of material similar to hospital oxygen tents), corrosion-proofing vital elements of the centrifugal air compressors (positive displacement blowers are not suitable), replacing the air compressor seals with oxygen compatible materials, gas piping, instrumentation, and an oxygen supply source. Stage baffles and multiple stage oxygen compressors and turbine-spargers are not used in this concept.

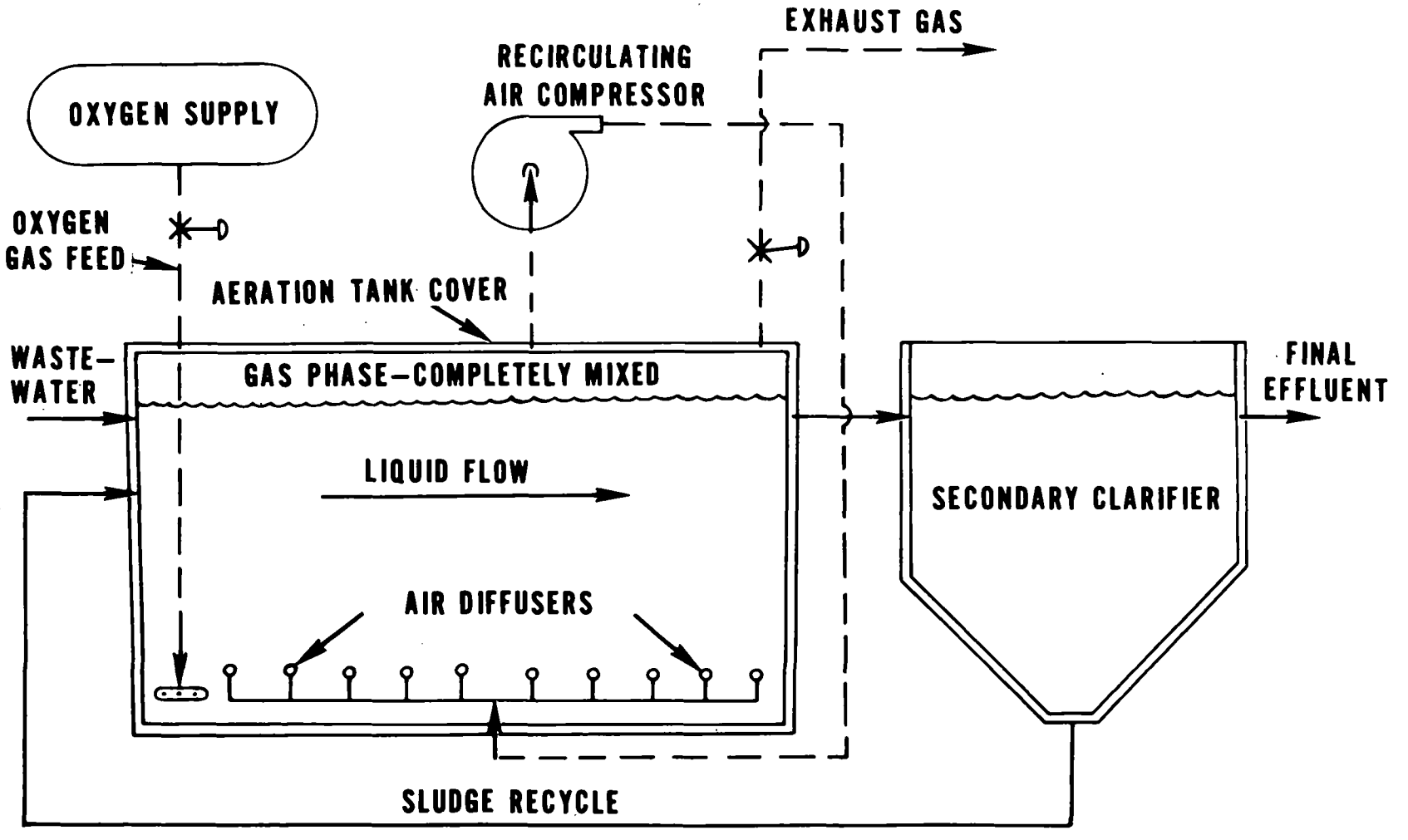
However, the utilization of existing equipment initially designed for conventional air aeration will result in less efficient gas transfer and higher specific power consumption than the multi-stage Union Carbide system. Because the Las Virgenes system is a single gas stage system, vapor phase gas will become completely mixed and assume a uniform composition identical to that of the exhaust gas. The available driving force for dissolving oxygen will be less than that available in the lead stages of the multi-stage concept.

At the conclusion of the project, the trade-offs (reduced capital costs at the sacrifice of increased operating costs) will be compared and their significance analyzed. Cordon International anticipates that its single stage system will produce equivalent treatment to the multi-stage system with volumetric organic loadings up to about 80% of those which can be handled by the multi-stage system.


TYPICAL RANGES
TOTAL TREATMENT COSTS
NEW PLANTS WITH PRIMARY SEDIMENTATION
FIGURE 3
AIR AERATION
OXYGEN AERATION
TOTAL TREATMENT COSTS - ¢/1000 GALLONS TREATED
PLANT SIZE-MGD
18
17
16
15
14
13
12
11
10
9
8
0
20
40
60
80
100

# FIGURE 4

## SCHEMATIC DIAGRAM OF DIFFUSED AIR AERATION SYSTEM MODIFIED TO RECIRCULATE OXYGEN GAS



GENERAL DISCUSSION - BLUE PLAINS

As seen in the on-going research projects, there are numerous applications for oxygen aeration processes, whether the feed is raw wastewater as at Batavia or primary effluent as in the Blue Plains study in Washington, D.C. Likewise, oxygen can be used to upgrade existing secondary treatment facilities as will be demonstrated in the Las Virgenes and Newtown Creek projects. With the different characteristics in operation and performance, the oxygen activated sludge system should be viewed and compared as an entire system composed of three interrelated subsystems; a biological reactor, a clarifier, and a solids handling system.

REACTOR

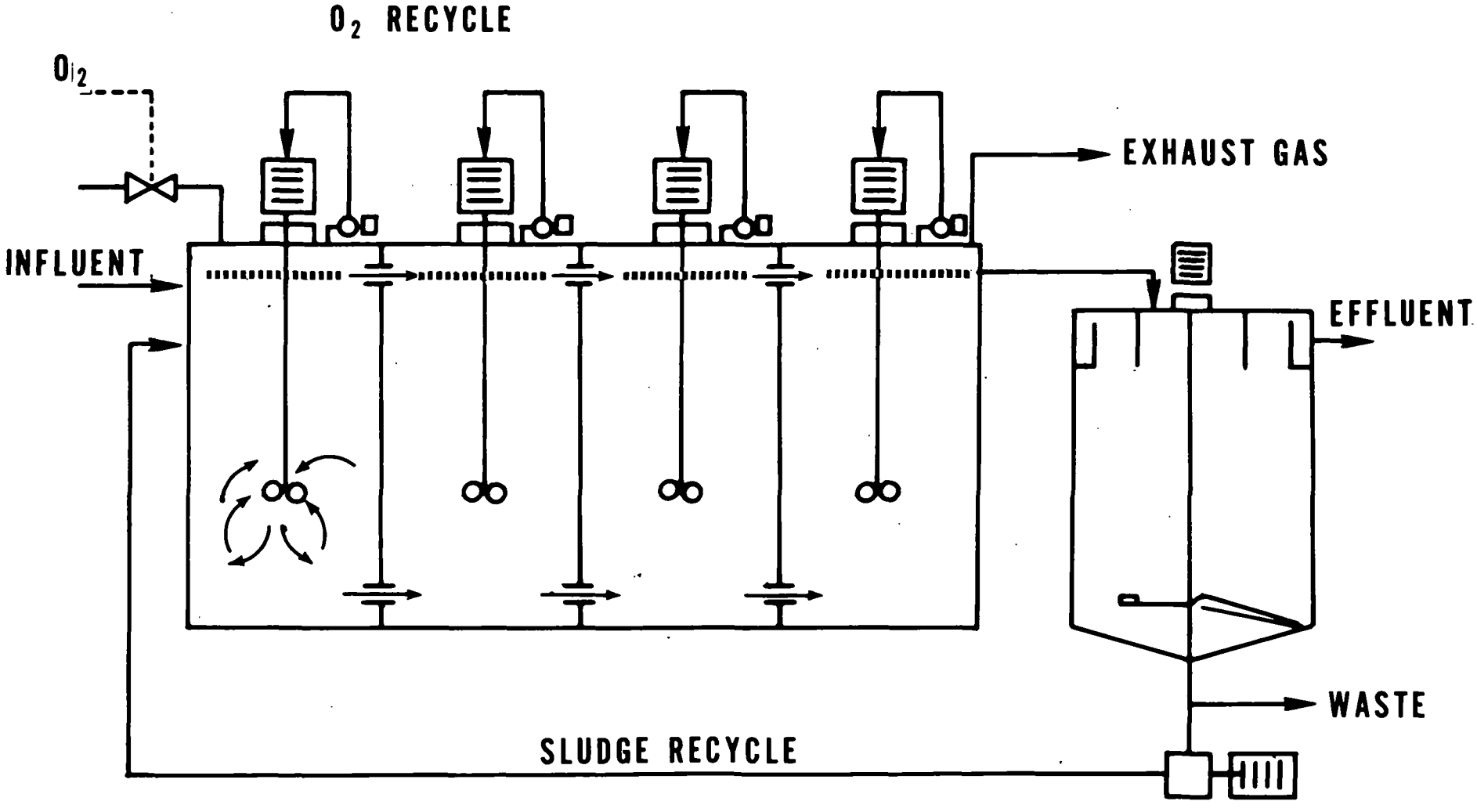
The first and most unique aspect of the system is the gas tight biological reactor shown in Figure 5. In the EPA/DC Pilot Plant, primary effluent from the District of Columbia's plant is fed to the oxygen reactor either on steady state flow or on a predetermined daily cycle (diurnal variation), normally with a 2.3:1 (45-105 gpm) daily flow variation.

Using all four available stages, the 8,100-gallon Blue Plains oxygen reactor provides 1.95 hours or detention time at the nominal influent flow of 100,000 gpd. At the peak daily flow, the detention time is 1.29 hours. Using three of the available four stages, the detention times are reduced to 1.50 hours and 1.00 hours, respectively, at the nominal and peak daily flows.

The reactor is sealed to prevent loss of oxygen and includes submerged hydraulic entrances and exits as well as simple water-sealed mixing equipment. Internal spray equipment using tap water is provided to suppress foam. Also, a partially submerged baffle plate before the internal exit trough retains

# FIGURE 5

## SCHEMATIC DIAGRAM OF BLUE PLAINS OXYGENATION SYSTEM



the foam until the baffle plate is raised to allow the foam build-up to escape. The reactor is staged to provide the proper tank geometry for efficient mixing and oxygen usage.

Efficient oxygen usage is achieved by co-current contacting of the mixed liquor and oxygen gas through the various stages. The addition of pure oxygen to the reactor is controlled by a pressure regulator. An inlet oxygen control valve actuated by a pressure regulator maintains the overhead gas at a selected pressure usually between 1" and 4" of water. Even with large instantaneous fluctuations in oxygen consumption, the oxygen control valve maintains the selected pressure. The overhead gas pressure is normally selected to maintain the oxygen concentration at approximately 50% in the exhaust gas from the last reactor stage. Pure oxygen is introduced to the first stage where the peak oxygen demand occurs. As the oxygen is used in biological metabolism, respirated carbon dioxide and stripped inert gases reduce the oxygen concentration in the overhead gas flowing co-currently with the mixed liquor through the succeeding stages. The successive decrease of both oxygen availability and oxygen demand produces efficient oxygen use before the residual gas is exhausted from the reactor.

Mixed liquor dissolved oxygen levels in the Blue Plains oxygen reactor are held between 4.0 and 8.0 mg/l by adjusting the recirculation rate of the oxygen gas within the individual stages. The compressor in each stage pumps the overhead gas through the rotating submerged turbine-sparger to provide efficient dispersion and mixing of the recirculated gas. The recirculation rate in each stage may be set either manually on the basis of the dissolved oxygen analysis or automatically using a control system with a dissolved oxygen sensor. The gas

recirculation rate in the first Blue Plains stage typically is 3-7 cfm and 1-2 cfm in each of the last three stages. Total recirculation requirements vary between 0.10 and 0.20 $ft^3$/gal. of flow.

With its high oxygen transfer capabilities (which are essentially independent of turbine mixing rates), the oxygen system is able to operate at higher MLSS concentrations. These factors enable the system to readily adsorb shock organic loads. Also, toxic shock loads can be better handled, much as in a totally mixed activated sludge system. Both types of systems initially expose the toxic substrate to a large mass of active solids and the resulting "biological inertia" buffers the toxicity.

On the District of Columbia wastewater, as seen in Figure 6, the volatile portion of the oxygen solids exhibit a much higher activity for the SRT range above 6 days than the District's step air aeration pilot process. The F/M ratio is the ratio of BOD applied to the mixed liquor volatile suspended solids (MLVSS) under aeration. Figure 6 indicates that a lower total volatile mass under aeration is required with oxygen than with air to obtain any given SRT above 6 days for a similar influent BOD. Thus, shorter detention times are possible with oxygen than with step aeration for similar MLSS concentrations to achieve any given SRT above 6 days. Further, at identical SRT's above 6 days, the oxygen system will produce less excess biological solids. The most probable reason for the increased activity is attributed to maintaining the mixed liquor dissolved oxygen between 4 and 8 mg/l. The independently controlled mixing also minimized sludge pockets, dead spots, and shearing of the floc particles. Mixed liquor entering the clarifier has a high dissolved oxygen content which permits a certain amount of aerobic metabolism in the clarifier and greatly reduces the time that the bio-mass is in an anoxic condition.

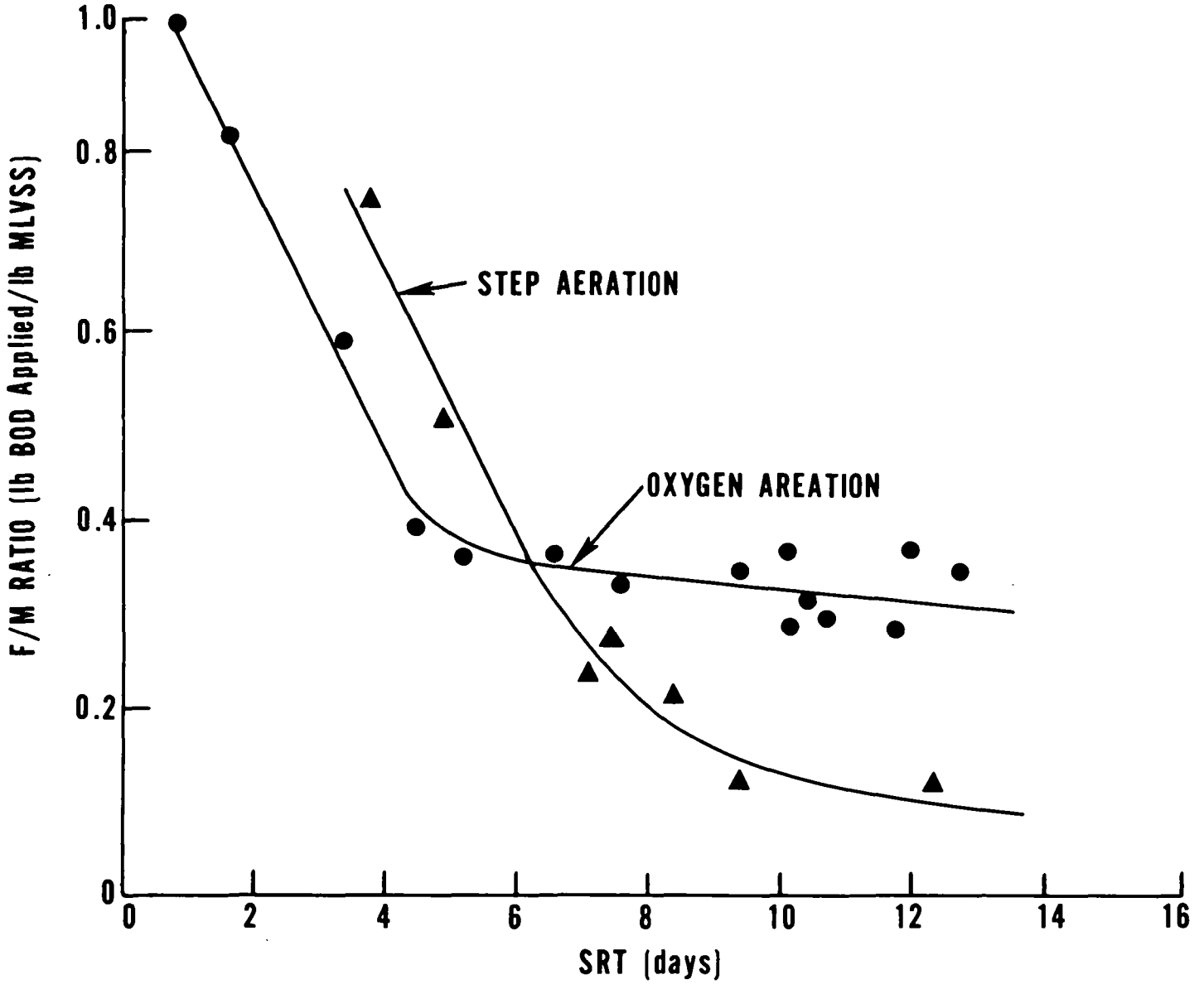



**FIGURE 6**
**BIOLOGICAL ACTIVITY RELATIONSHIPS - BLUE PLAINS**

The total production of solids in the oxygen system (Figure 7) per pound of BOD added, including underflow waste and effluent solids, is inversely related to the solids retention time (SRT) above an SRT of 1.3 days. The solids production with the oxygen aeration system was significantly lower than in the conventional step aeration pilot system above an SRT of 6 days (or than in a high rate modified air aeration pilot process tested by the District). Indeed, the total solids production decreased from 0.65 pounds of excess solids per pound of BOD added at an SRT of 6 days to 0.35 pounds of excess solids per pound of BOD added at an SRT of 13 days with only a 33% increase in volatile solids concentration at the higher SRT. The parallel conventional system, operated as step aeration or contact stabilization, exhibited increased solids production through an SRT of 9.5 days with a peak solids production of approximately 1 pound of excess solids per pound of BOD added. However, an approximate four-fold increase in volatile solids was required to raise the SRT from 6 to 13 days in this system and to begin to achieve reduced solids production. The modified aeration system, in log growth rate biology, produced solids at the rate of 1.0 to 2.0 pounds of excess solids per pound of BOD added at its operating SRT of less than 1 day (115 mg/l of alum added for P removal).

The reduction of BOD in the reactor was excellent. With an influent BOD up to 130 mg/l, a wide range of detention times from 1.5 to 2.5 hours, and SRT's that varied from 13 to as low as 2 days, the effluent soluble BOD was consistently less than 5 mg/l as described in Table 1. This indicates virtually complete insolubilization of the BOD in the primary effluent. Thus, BOD removal on the D.C. oxygen system is a function of clarification.

The oxygen mixed liquor was similar visually to the micro-organisms in conventional activated sludge (Figure 8). The mixed liquor biota was normally

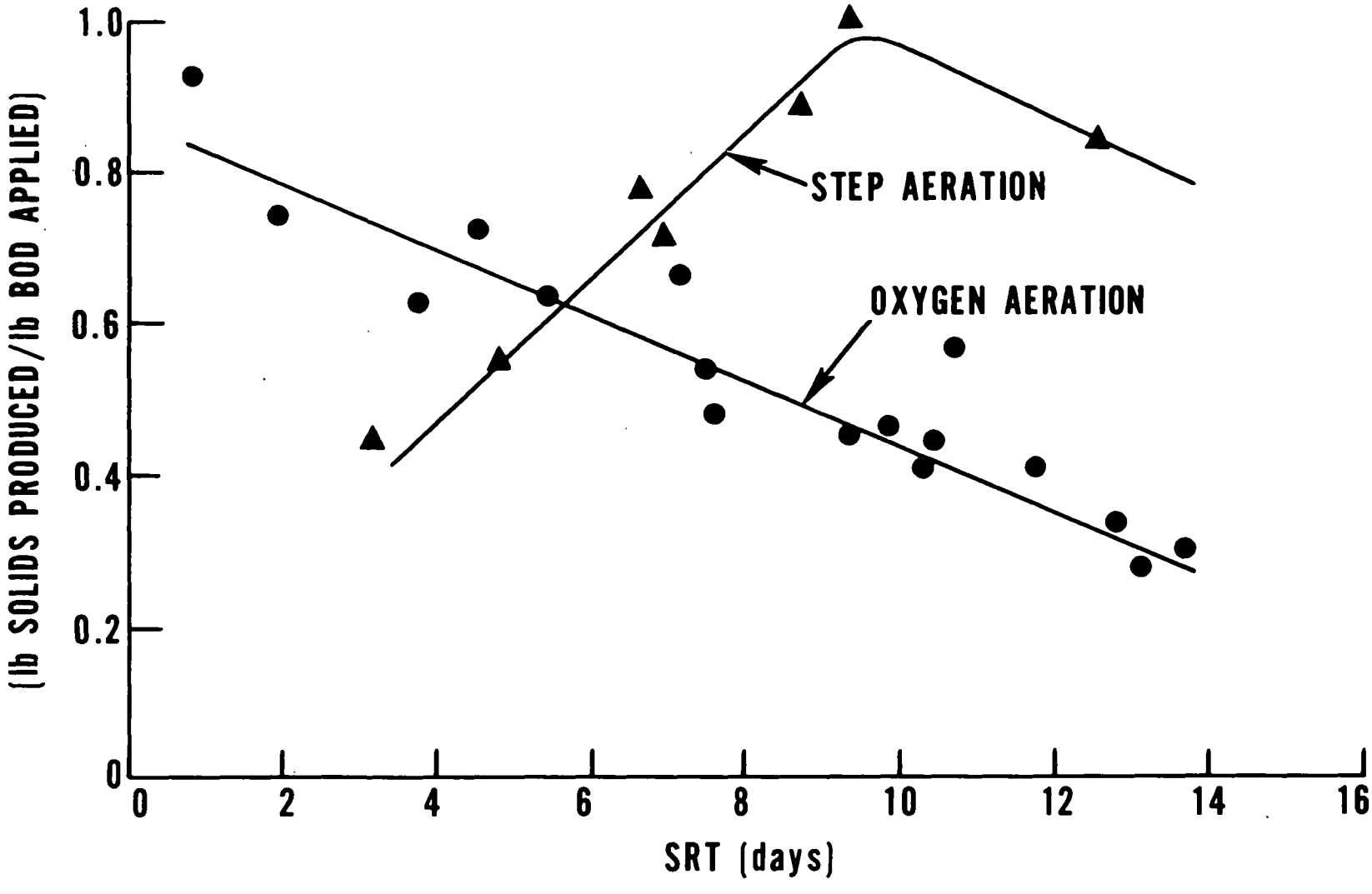



FIGURE 7

EXCESS BIOLOGICAL SLUDGE PRODUCTION - BLUE PLAINS

TABLE 1

ORGANIC REMOVAL - BLUE PLAINS

| Operating Period<br>Month<br>Dates | 1<br>June<br>12-30 | 2<br>July | 3<br>August<br>1-25 | 4<br>September | 5<br>October<br>3-11 | 6<br>November<br>10-30 | 7<br>January<br>1-16 | 8<br>January<br>17-31 |
|---|---|---|---|---|---|---|---|---|
| Primary Effluent BOD (mg/l) | 89 | 87 | 89 | 106 | 116 | 131 | 124 | 134 |
| Final Effluent BOD (mg/l) | 18 | 19 | 12 | 13 | 14 | 27 | 11 | 32 |
| Final Effluent Soluble BOD (mg/l) | - | - | 2 | 2 | 3 | 3 | 3 | 3 |
| Primary Effluent COD (mg/l) | 250 | 244 | 245 | 252 | 284 | 275 | 250 | 256 |
| Final Effluent COD (mg/l) | 45 | 70 | 49 | 51 | 51 | 63 | 59 | 99 |
| Primary Effluent TOC (mg/l) | 75 | 65 | 77 | 100 | 106 | 91 | 83 | 87 |
| Final Effluent TOC (mg/l) | 14 | 24 | 15 | 15 | 15 | 21 | 21 | 26 |
| Primary Effluent Suspended Solids (mg/l) | 113 | 101 | 102 | 107 | 120 | 92 | 98 | 100 |
| Final Effluent Suspended Solids (mg/l) | 36 | 53 | 28 | 24 | 35 | 56 | 24 | 58 |

TABLE 1 (CONTINUED)

ORGANIC REMOVAL - BLUE PLAINS

| 9 March 1-18 | 10 April | 11 May | 12 June | 13 July | 14 August | 15 September | 16 October | 17 November | 18 December 1-21 | 19 December 22-31 |
|---|---|---|---|---|---|---|---|---|---|---|
| 121 | 107 | 140 | 110 | 129 | 110 | 149 | 120 | 125 | 125 | 135 |
| 27 | 10 | 7 | 8 | 14 | 15 | 15 | 14 | 20 | 20 | 18 |
| - | 4 | 4 | - | - | - | 4 | 4 | 5 | 5 | 5 |
| 251 | 267 | 278 | 238 | 235 | 219 | 239 | 224 | 244 | 238 | 236 |
| 76 | 48 | 51 | 45 | 35 | 32 | 35 | 37 | 54 | 59 | 53 |
| 88 | 81 | 92 | 74 | 78 | 69 | 79 | 69 | 75 | 76 | 91 |
| 22 | 17 | 18 | 18 | 14 | 13 | 15 | 14 | 17 | 19 | 19 |
| 104 | 83 | 120 | 100 | 103 | 97 | 95 | 81 | 90 | 95 | 95 |
| 49 | 18 | 12 | 13 | 11 | 16 | 15 | 15 | 23 | 23 | 18 |

FIGURE 8

very well bioflocculated with active stalked cilates growing on the bacterial mass. Zooflagellates and free swimming cilates, although few in number, remained adjacent to or within the flocculated particles. Several varieties of large active rotifers were present in abundance. A few nematodes existed in the sludge. Normally, filamentous growth was not apparent. There was almost complete absence of fragmented debris or unflocculated bacteria between the discrete particles. In the SRT range less than 5 days, both the oxygen activated sludge and the conventional aeration systems exhibited filamentous growth on the District of Columbia wastewater. Filamentous growth did not occur during operation above an SRT of 5 days. Normally, when encountering filamentous growth for a few days, reducing system influent flow to increase the SRT reestablished a filamentous free sludge in several days. However, after extended periods of operation with filamentous growth, the Sphaerotilis became firmly entrenched and could not be quickly purged from the system by flow reduction techniques. Hydrogen peroxide added to the recycle in two 24-hour periods approximately a week apart at dosages of 200 mg/l (based on influent flow) was then required to purge the system of filamentous growth.

The recirculation of respirated carbon dioxide within the oxygen reactor stages lowers the wastewater pH from 7.0 - 6.8 in the first stage and to 6.4 - 6.1 in the final stage. With an average system pH of approximately 6.5, the oxygen process more slowly establishes a nitrifying population than the step aeration activated sludge system operated at a pH of 7.0 to 7.4. However, during the warmer months when the solids wasting is reduced to a level where the nitrifying organisms propagate faster than they are removed, the Nitrosomonas and Nitrobacter populations increase and substantial nitrification occurrs in the oxygen system. Nitrogen removal across the oxygen system during periods

of high nitrification and partial denitrification is as high as 39-40%. Nitrogen removal decreases to a low of 9-10% during periods without nitrification.

As in the parallel step aeration process, nitrification in the oxygen system begins to decrease in the Fall and becomes virtually nil during the Winter. At wastewater temperatures of about 63°F, 5 mg/l of $NO_3$-N is still produced with an SRT of 9.0 days.

Without alum addition, phosphorus is removed from the oxygen system through metabolic uptake and by wasting of excess solids; thus, the removals vary with the metabolism of the mixed liquor. At high SRT's (highly endogenous metabolism), total phosphorus removal averages only about 15%. At lower SRT's with less endogenous respiration, phosphorus removals increase to 20%.

With alum addition, phosphorus removal in the oxygen system increases as the alum weight ratio ($Al^{+++}$/P) increases. During experiments conducted in the Fall of 1970, for a dosage equal to an $Al^{+++}$/P ratio of 1.4/1, 80% of the phosphorus was removed to an average residual of 1.8 mg/l as P and only a slight decrease in wastewater alkalinity and pH occurred. The filtered effluent (though 0.45$\mu$) contained an average of 1.6 mg/l of soluble P. When the dosage was increased to a ratio of 1.85/1 ($Al^{+++}$/P), the residual total and soluble phosphorus decreased to 0.62 and 0.53 mg/l as P, respectively. At this higher dosage, however, the buffering capacity of the oxygen mixed liquor was further reduced and the average pH decreased from 6.5 to 6.0. The oxygen biomass dispersed, necessitating termination of the alum addition to allow the mixed liquor to recover. In areas with low alkalinity wastewaters, additional alkalinity in the form of lime or caustic may be required to control pH at a level which will prevent floc dispersion. This pH adjustment may be necessary in either air or oxygen systems but is more likely in an oxygen system because of the increased dissolved $CO_2$ content of the mixed liquor. The addition of alum and precipitation of $Al(PO_4)$ and $Al(OH)_3$ increases the inert solids carried in the system and adequate clarification for the higher solids concentration must be provided.

Consistently throughout the operation, vented gas from the fourth oxygen reactor stage has been less than 10% of the input oxygen volume. The vented stream is roughly 50% oxygen. Based upon the influent and exhausted oxygen concentrations, the net utilization of oxygen in the process is about 95%. The accountable oxygen consumption consisting of COD removal, nitrification demand, exhaust gas, and effluent dissolved oxygen is summarized in Table 2. The COD removed was calculated by substracting the COD in the underflow waste solids and that in the process effluent from the primary effluent COD. With increasing SRT, additional oxygen is required for endogneous respiration. Likewise during periods with nitrification, additional oxygen is required.

Pertinent reactor variables and operating conditions are summarized in Table 3 for approximately 1-1/2 years of operation on the Blue Plains oxygen aeration pilot plant.

CLARIFICATION

The second important aspect in the oxygen aeration system is liquid/solids separation. At the EPA/DC Pilot Plant, as at Batavia, gravity clarification is employed. As mentioned before, soluble residual BOD in the effluent averaged less than 5 mg/l in the test periods indicating virtually complete BOD insolubilization. Thus, most of the residual BOD in the Blue Plains oxygen system effluent is associated with suspended solids. Overall removal of suspended solids and BOD is a function of clarification efficiency.

Clarifier efficiency is in turn a function of the basic settling characteristics of the solids as well as of the actual design and operation of the clarifier. With the normally higher mixed liquor concentrations used in the oxygen aeration process, design criteria for both clarifiers (i.e., overflow rates and volume) and thickeners (i.e. solids loading - $lb/ft^2/day$) should be considered. The Ten State Standards suggest that conventional activated sludge

TABLE 2 - OXYGEN USAGE - BLUE PLAINS

| Operating Period<br>Month<br>Dates | 1<br>June<br>12-30 | 2<br>July | 3<br>August<br>1-25 | 4<br>September | 5<br>October<br>3-21 | 6<br>November<br>10-30 | 7<br>January<br>1-16 | 8<br>January<br>17-31 |
|---|---|---|---|---|---|---|---|---|
| Primary Effluent COD (lb/million gal.) | 2080 | 2030 | 2040 | 2090 | 2370 | 2290 | 2080 | 2140 |
| Final Effluent COD (lb/million gal.) | 375 | 584 | 408 | 430 | 425 | 525 | 488 | 826 |
| Waste Sludge COD (lb/million gal.) | 188+ | 42+ | 150+ | 217+ | 630+ | 247 | 233 | 336 |
| COD Removed from System (lb/million gal.) | 1517 | 1404 | 1482 | 1443 | 1315 | 1518 | 1359 | 978 |
| Nitrate Nitrogen Demand (lb/million gal.) | 14 | 69 | 128 | 160 | 100 | 18 | 9 | 14 |
| Exhaust Oxygen (lb/million gal.) | 85 | 54 | 75 | 85* | 87 | 85* | 65 | 65 |
| Final Effluent D.O. (lb/million gal.) | 10* | 10* | 10 | 10 | 10 | 10* | 25 | 25 |
| Total | 1626 | 1537 | 1695 | 1698 | 1512 | 1631 | 1458 | 1082 |
| Oxygen Supplied (lb/million gal.) | 1750 | 1825 | 1775 | 1900 | 1650 | ± | 1700** | 1800** |

+ COD = 1.4 volatile solids
* Estimate
± Inlet meter malfunctioned
** Increase sampling and greater losses of $O_2$ through sample ports

TABLE 2 - OXYGEN USAGE - BLUE PLAINS (CONT'D)

| 9 March 1-18 | 10 April | 11 May | 12 June | 13 July | 14 August | 15 September | 16 October | 17 November | 18 December 1-21 | 19 December 22-31 |
|---|---|---|---|---|---|---|---|---|---|---|
| 2050 | 2180 | 2500 | 1750 | 1940 | 1830 | 1920 | 1880 | 1950 | 1990 | 1970 |
| 630 | 380 | 450 | 340 | 270 | 260 | 280 | 310 | 440 | 490 | 440 |
| 160 | 700 | 890 | 80 | 270 | 270 | 400 | 390 | 220 | 230 | 300 |
| 1260 | 1100 | 1160 | 1330 | 1400 | 1300 | 1240 | 1180 | 1290 | 1270 | 1230 |
| 0 | 0 | 0 | 0 | 60 | 70 | 270 | 200 | 220 | 190 | 70 |
| 160 | 130 | 40 | 80 | 300 | 260 | 200* | 200* | 200* | 200* | 150* |
| 30 | 40 | 50 | 60 | 60 | 40 | 40 | 30 | 40 | 40 | 40 |
| 1450 | 1270 | 1250 | 1470 | 1820 | 1670 | 1750 | 1610 | 1750 | 1700 | 1490 |
| 1450 | 1300 | 1260 | 1600 | 2200 | 1690 | ± | ± | 1740 | 2000 | 1450 |

TABLE 3

REACTOR VARIABLES AND OPERATING CONDITIONS

BLUE PLAINS

| Operating Period<br>Month<br>Dates | 1<br>June<br>12-30 | 2<br>July | 3<br>August<br>1-25 | 4<br>September | 5<br>October<br>3-21 | 6<br>November<br>10-30 | 7<br>January<br>1-16 | 8<br>January<br>17-31 |
|---|---|---|---|---|---|---|---|---|
| Flow Rate (gpm) | 50-55 | 80 | 80 | 70+ | 70+* | 70+ | 53 | 53 |
| Aeration Time (hr) | 2.00 | 1.66 | 1.66 | 1.95 | 1.95 | 1.95 | 2.50 | 2.50 |
| Recycle Rate | 50% | 50% | 42% | 32% | 38% | 37% | 77% | 80% |
| MLSS (mg/l) | 4140 | 5180 | 5250 | 6000 | 8120 | 6350 | 5300 | 3940 |
| MLVSS (%) | 74% | 70% | 73% | 78% | 67% | 73% | 80% | 81% |
| SRT (days) | 7.7 | 7.3 | 11.8 | 10.7 | 5.5 | 5.5 | 13.0 | 4.7 |
| F/M (lb BOD/day/lb MLVSS) | 0.333 | 0.342 | 0.296 | 0.304 | 0.283 | 0.355 | 0.275 | 0.392 |
| Volumetric Loading (lb BOD/day/1,000 $ft^3$) | 57 | 80 | 80 | 96 | 106 | 108 | 89 | 80 |
| Mixer Power $\left(\frac{k.w.-hr}{1,000\ gal.}\right)$ ** | 1.27 | 1.19 | 0.98 | 0.92 | 1.00 | 1.00 | 1.18 | 1.42 |
| Compressor Power $\left(\frac{k.w.-hr}{1,000\ gal.}\right)$ ** | 0.39 | 0.28 | 0.39 | 0.41 | 0.35 | 0.26 | 0.34 | 0.32 |
| Temperature (°F) | 74-80 | 78-84 | 82-85 | 79-83 | 70-79 | 66-69 | 58-60 | 58-60 |

* Alum addition
\+ 2.3:1 diurnal variation
** Pilot plant equipment efficiency was not determined

TABLE 3 (CONTINUED)

REACTOR VARIABLES AND OPERATING CONDITIONS - BLUE PLAINS

| 9 March 1-18 | 10 April | 11 May | 12 June | 13 July | 14 August | 15 September | 16 October | 17 November | 18 December 1-21 | 19 December 22-31 |
|---|---|---|---|---|---|---|---|---|---|---|
| 60-70 | 31-67 | 60 | 30-70 | $70^+$ | $70^+$ | $70^+$ | $70^+$ | $70^+$ | $70^+$ | $70^+$ |
| 2.15 | 3.30-1.55 | 1.70 | 3.70-1.50 | 1.5 | 1.5 | 1.5 | 1.5 | 2.0 | 2.0 | 2.0 |
| 50-60% | 90-60% | 65% | 100-50% | 50% | 50% | 46% | 36% | 30% | 25% | 25% |
| 3070 | 2710 | 2750 | 4000 | 6600 | 7500 | 7400 | 6000 | 4600 | 4400 | 4200 |
| 77% | 81% | 78% | 73% | 70% | 70% | 72% | 73% | 78% | 80% | 81% |
| 3.7 | 1.3-4.0 | 2.0 | 13.0 | 12.6 | 10.0 | 7.5 | 9.5 | 9.8 | 9.0 | 6.5 |
| 0.580 | 0.30-1.00 | 0.970 | 0.400 | 0.430 | 0.32 | 0.39 | 0.31 | 0.39 | 0.40 | 0.50 |
| 90 | 98 | 157 | 95 | 160 | 131 | 185 | 146 | 111 | 111 | 122 |
| - | - | - | - | - | - | - | - | - | - | - |
| - | - | - | - | - | - | - | - | - | - | - |
| 60-62 | 62-65 | 65-71 | 70-77 | 77-80 | 77-81 | 76-81 | 76-71 | 71-65 | 65-63 | 63-61 |

clarifiers be designed for average overflow rates of 800 gpd/ft$^2$. The Water Pollution Control Federation Manual of Practice (1959) suggests that the solids loading be held below a peak of 30 lb/ft$^2$/day. Overflow rates and solids loading criteria should be better defined for high solids systems such as oxygen aeration. An internal EPA study is presently underway on several oxygen aeration pilot facilities (with the cooperation of industry) to further investigate sludge settling parameters.

At this point, what we have learned to date will be discussed. The basic settling characteristics of the mixed liquor typically have been found to be a function of:

1. Concentration of the mixed liquor
2. Particle shape
3. Particle density
4. Seasonal variation
    a. Physical changes in water density and viscosity with temperature
    b. Metabolic changes with temperature
    c. Seasonal loading variation

As seen in Figure 9, the log of the initial settling rate is a function of the log of the solids concentration. This sample curve illustrates that two relationships exist. The first at lower MLSS levels corresponds to free particle settling and is characterized by the absence of an initial discrete subsiding interface and a zone of homogenous settling solids. The second at higher MLSS levels has both an initial discrete interface and a zone of homogenous settling particles (zone settling). Thus, the sizing of a clarifier is a function of the MLSS concentration and must be coordinated with the reactor (and the sludge handling facilities) to achieve the desired biological capabilities of the system.

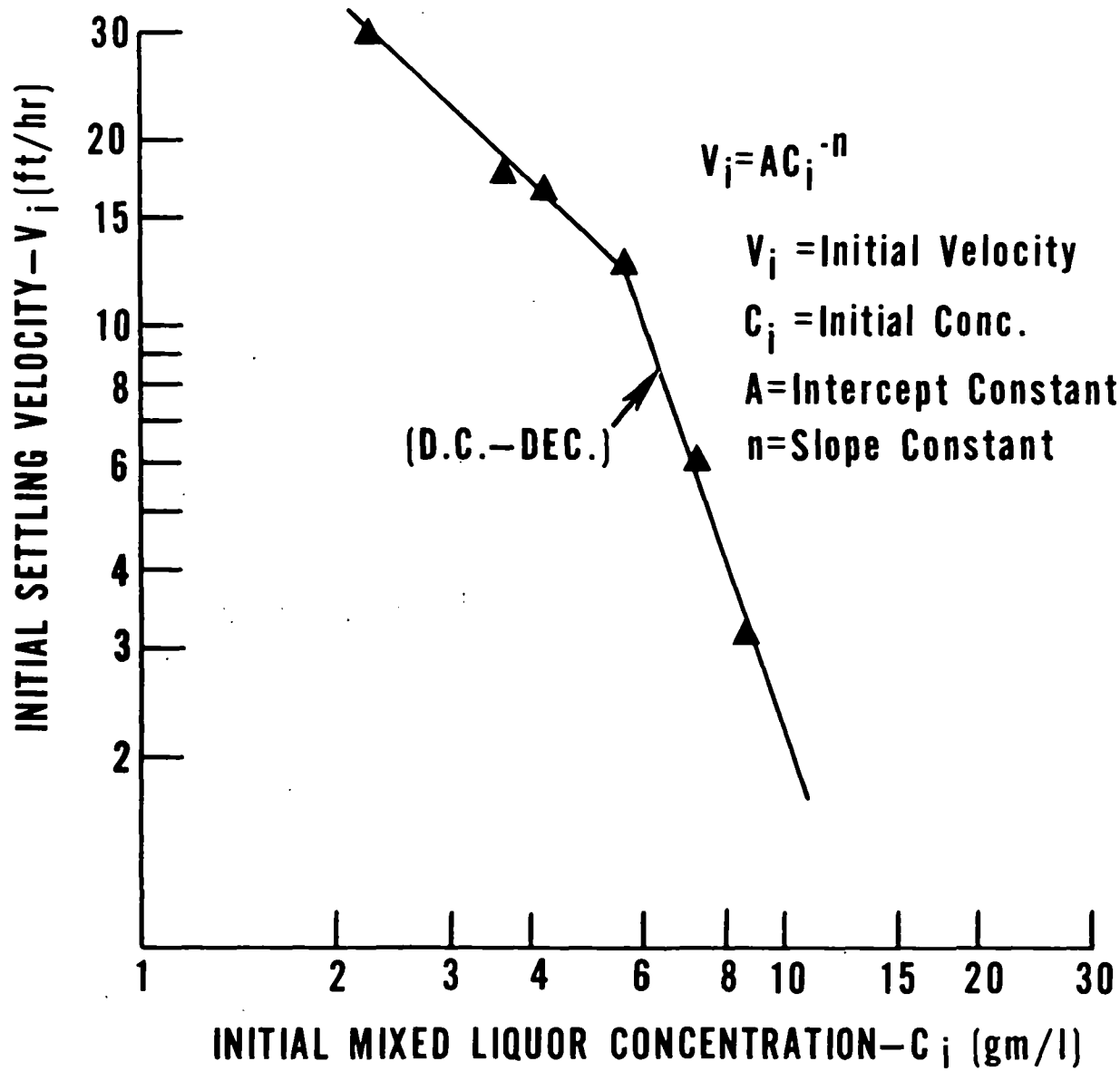



**FIGURE 9**

**INITIAL SLUDGE SETTLING VELOCITY PROFILE FOR BROAD MIXED LIQUOR CONCENTRATION RANGE - BLUE PLAINS**

Another important factor is the particle shape. Normally, as shown in Figure 8, the oxygen mixed liquor particles have rounded shapes. However, if filamentous growth exists, as experienced below an SRT of 5 days in D.C. (air and oxygen), both settling rates and compaction deteriorate. The range, if any, that filamentous growth appears is unique to each location and should be defined for that location. The presence of industrial fibers would have an effect similar to filamentous growth on mixed liquor settling characteristics.

Still another important factor in the basic settling characteristics is the density of the particles in relationship to that of the water. It is the difference in density that is the driving force for settling. The VSS/TSS ratio (or volatile %) is one relative indication of density. There are several ways to improve the particle density. One is to feed raw wastewater instead of primary effluent to the oxygen aeration system, thus incorporating the normally denser particles captured in primary sedimentation into the biomass, such as occurred at Batavia. Again, the sizing of the reactor oxygen supply, etc. must be compatible with the increased organic loading. In Washington, heavy rains and unusually high flows wash silt and clay into the sewer system. These materials subsequently become incorporated in the mixed liquor solids and have increased sludge settling rates 30% to 60%. In like fashions, operation under different biological conditions can alter sludge settling characteristics.

Another unique method of increasing the density of the sludge was evaluated at the EPA/DC Pilot Plant by altering the method of clarifier operation. Two major methods of clarifier operation are possible. One is to use the blanket as a filter and the other is to permit classification of the settling solids. The first method can be accomplished in two different ways: (1) By providing sufficient depth to the clarifier such that the mixed liquor passes up through the clarifier blanket (which acts as a filter). (2) By carrying high MLSS

concentrations (usually above 4500 mg/l in D.C.) such that the particles settle in a subsidence (zone) settling pattern with discrete interfaces existing between the homogeneous subsiding particles and the decant. In the subsidence zone, the relatively uniform concentration of particles are nearly homogeneously mixed by the countercurrent turbulence produced by water passing around the solids. The homogeneous subsiding blanket does not allow classification of individual particles because the settling blanket acts as a filter.

At Blue Plains, with MLSS concentrations below 4500 mg/l, subsidence (zone) settling does not occur during the initial portion of settling. This provides for a second method of clarifier operation where classification of the discrete settling particles can occur if the mixed liquor is fed above the clarifier blanket level. The lighter or unsettlable particles, thus, can be purged from the system. The effluent suspended solids accordingly increased from 15 mg/l to 25 mg/l during this method of operation with a corresponding increase in effluent BOD.

Seasonal variations also affect sludge settling characteristics in oxygen as well as in air systems. These variations become critical as the MLSS of the mixed liquor increases. The pure physical changes in the wastewater density and viscosity contribute to slower settling rates as the wastewater temperature decreases. As the density of the water increases, the driving force for settling (which is the difference in density between water and the settling particles) decreases for a similar particle density. The drag force, viscosity, also increases with decreasing temperature (~25% from 80°F to 55°F) again contributing to slower settling rates in colder waters. Figure 10 shows a series of liter batch settling tests conducted in June, 1971 by only altering the temperature of the mixed liquor. As expected, the colder samples settled slower. In Figure 11, the batch flux (concentration multiplied by settling velocity) or the solids loading in

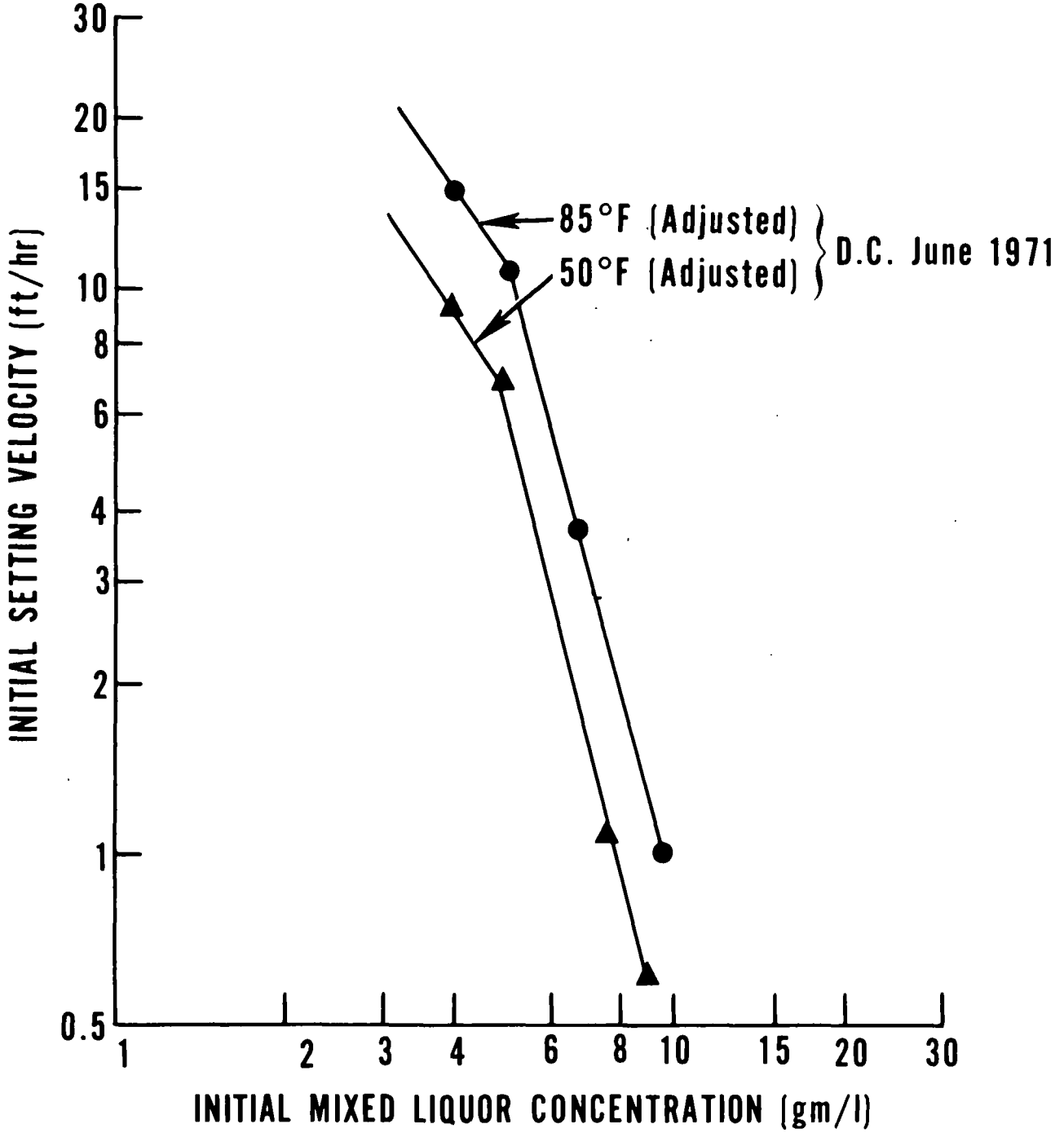



**FIGURE 10**

**EFFECT OF ADJUSTED VS. ACCLIMATED WASTEWATER TEMPERATURES ON SLUDGE SETTLING RATES - BLUE PLAINS**

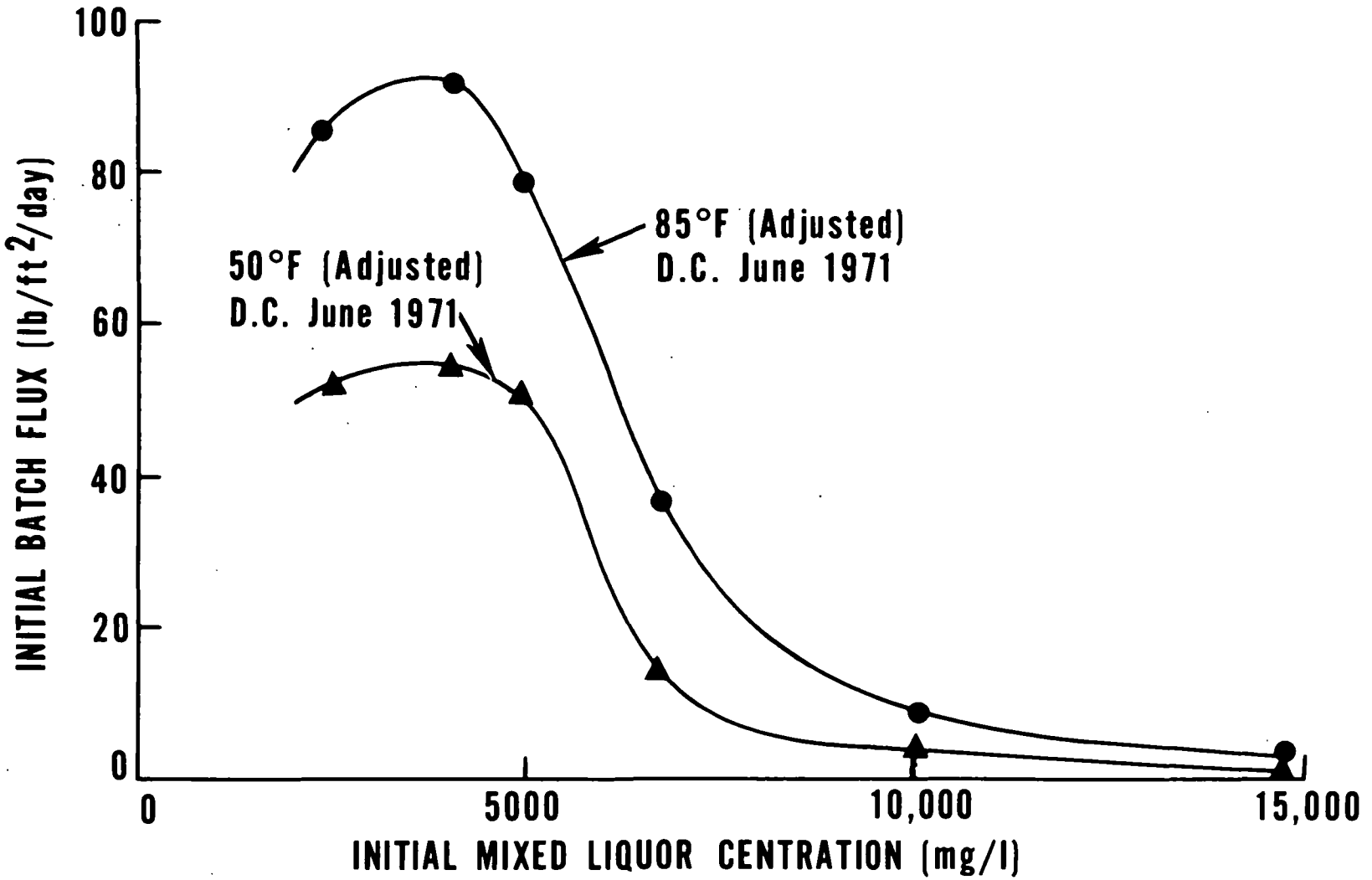



FIGURE 11

EFFECT OF WASTEWATER TEMPERATURE ON INITIAL BATCH FLUX - BLUE PLAINS

$lb/ft^2/day$ is shown for the previous tests. Solids loading is often used in thickener design. Again, the effect of wastewater temperature is evident.

Besides the physical changes caused by seasonal variation, another factor which must be considered is the metabolic change brought about by changing wastewater temperature. Figure 12 shows that the settling characteristics of oxygen mixed liquor change seasonably at D.C. At similar SRT's, the initial settling rate in a 1 liter graduated cylinder test decreased from approximately 10 ft/hr to 7 ft/hr at a concentration of 6000 mg/l as the temperature changed from 81°F to 71°F. The clarifier was being operated to capture unsettleable particles at this time. In Figure 13, the clarifier was operated to purge the unsettleable particles; but, again the solids showed a decreasing initial settling rate with decreasing wastewater temperature for a similar biology. The initial settling rate in the 1 liter test decreased from 14 ft/hr to 9 ft/hr at 4500 mg/l as the temperature decreased from 70°F to 63°F. Similar patterns of decreasing settling rates with decreasing wastewater temperature have been observed in nitrifying and denitrifying mixed liquors also.

Clarifier operation and design are equally important to the basic settling characteristics of the solids in gravity clarification. Besides selecting an overflow rate compatible with reactor sizing, the depth and method of clarifier feed are important as discussed earlier for either high solids capture or solids classification. Other important design considerations are the volume or detention time of the clarifier and recycle rate. At Blue Plains, the oxygen system underflow solids concentration varies between 1.0% and 1.4% with an average clarifier detention time of 1.9 hours. With 2.8 hours average detention time, the underflow solids concentration rises to 2.0%-2.4% with similar recycle rates. The sludge recycle rate is then determined after an F/M ratio is established for the reactor

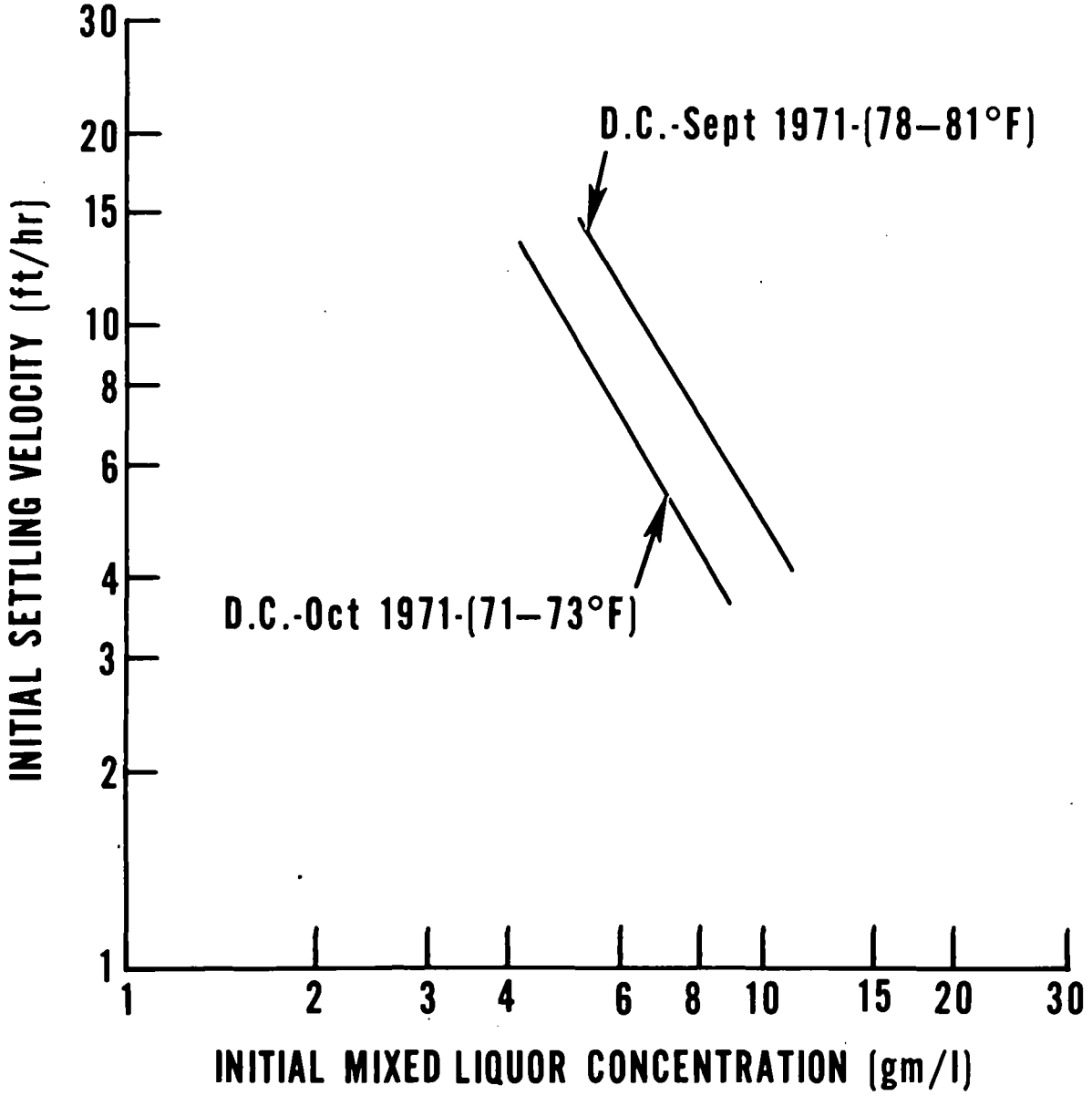



**FIGURE 12**

**EFFECT OF WASTEWATER TEMPERATURE ON OXYGENATED SLUDGE SETTLING RATES - BLUE PLAINS (SEPT.-OCT.,1971)**

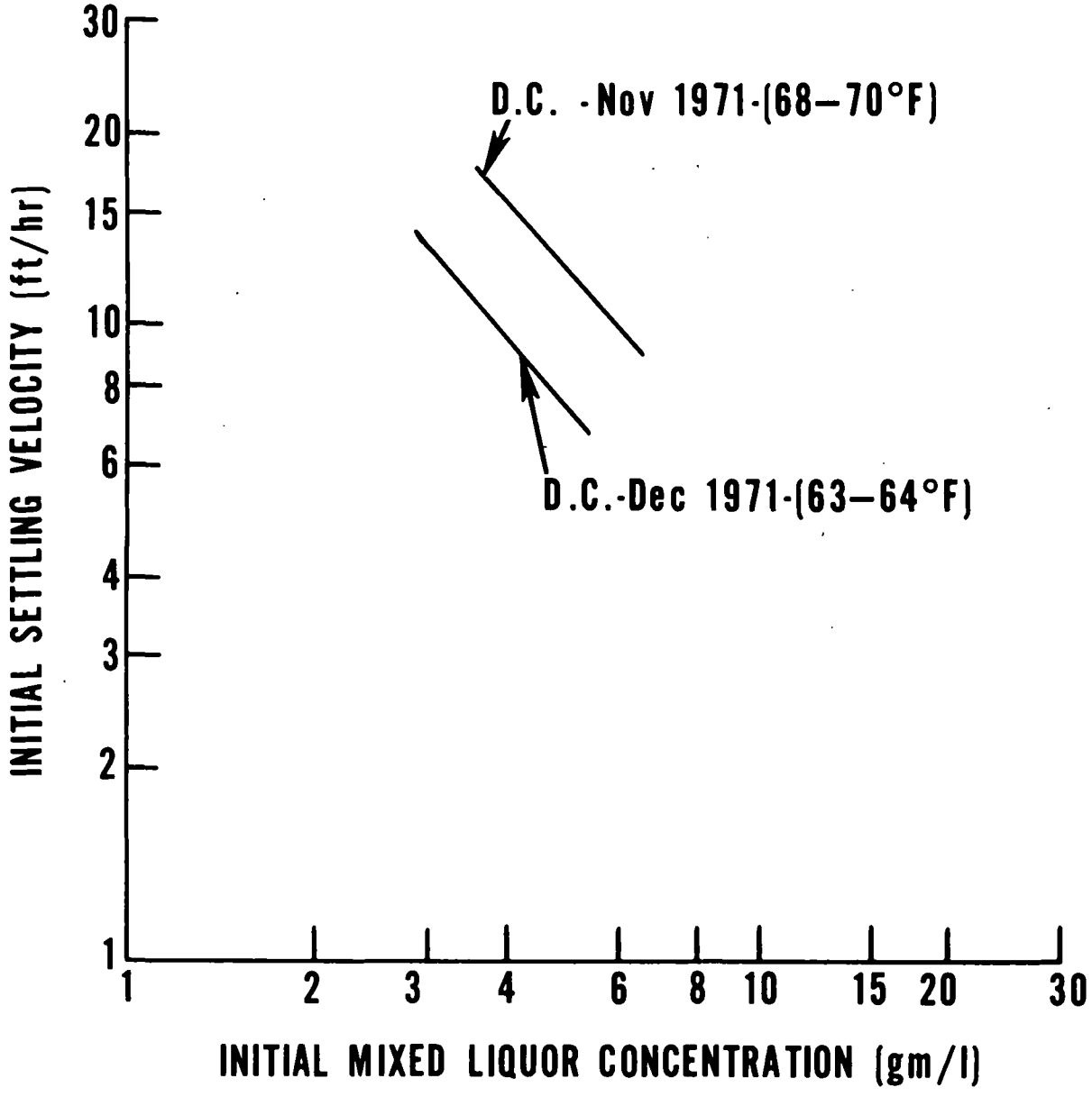



**FIGURE 13**

**EFFECT OF WASTEWATER TEMPERATURE ON OXYGENATED SLUDGE SETTLING RATES - BLUE PLAINS (NOV.-DEC.,1971)**

and the underflow concentration from the clarifier is likewise established.

Inventory solids are another consideration in clarifier operation. The total solids inventory is a result of both the build-up of solids in the blanket level and the solids actually in the transition (or settling) process. A simple increase in the wasting rate will reduce the blanket level if the increase is caused by the build-up of solids. However, as the initial settling rates decreased, as in Washington, for a given MLSS concentration, the sludge inventory also increases in the clarifier as more solids are in transition (or settling). In this case, increased wasting rates will not reduce the solids inventory in the clarifier without thinning the MLSS. It appears that the rising blanket can be a result of slower settling solids in the transition zone rather than a backlog of solids due to inadequate wasting.

In the Summer of 1970 at an MLSS concentration in excess of 8000 mg/l, peak clarifier overflow rates of 1940 gpd/ft$^2$ were observed on the Blue Plains oxygen system as shown in Table 4. During the 1970-71 Winter, the peak sustained overflow rates which could be maintained without the blanket coming over the weirs was 975 gpd/ft$^2$ at MLSS concentrations that varied from 3900 to 5300 mg/l. The causative agents which reduced the allowable overflow rates for satisfactory operation from Summer to Winter were undoubtedly a combination of all the above mentioned factors, not the least of which was the decreased wastewater temperature.

TABLE 4

CLARIFIER VARIABLES AND OPERATING CONDITIONS

BLUE PLAINS

| Operating Period<br>Month<br>Dates | 1<br>June<br>12-30 | 2<br>July | 3<br>August<br>1-25 | 4<br>September | 5<br>October<br>3-21 | 6<br>November<br>10-30 | 7<br>January<br>1-16 | 8<br>January<br>17-31 |
|---|---|---|---|---|---|---|---|---|
| Average Overflow Rate ($gpd/ft^2$) | | | | | | | | |
| At Surface | - + | - + | - + | 1280± | 1280± | 1280± | 975± | 975± |
| Above Feed Skirt | 750 | 1210 | 1210 | 1050 | 1050 | 1050 | 800 | 800 |
| Below Feed Skirt | 670 | 1075 | 1075 | 940 | 940 | 940 | 710 | 710 |
| Peak Overflow Rate ($gpd/ft^2$) | | | | | | | | |
| At Surface | - + | - + | - + | 1940 | 1940 | 1940 | 975 | 975 |
| Above Feed Skirt | 750 | 1210 | 1210 | 1580 | 1580 | 1580 | 800 | 800 |
| Below Feed Skirt | 670 | 1075 | 1075 | 1410 | 1410 | 1410 | 710 | 710 |
| Average Solids Loading ($lb/day/ft^2$) | 37 | 75 | 58 | 61 | 88 | 68 | 55 | 42 |
| SVI | 80 | 48 | 50 | 42 | 33 | 48 | 60 | 73 |
| Underflow Solids | | | | | | | | |
| % Dry Solids | 1.16 | 1.34 | 1.27 | 1.40 | 2.14 | 1.40 | 1.08 | 1.00 |
| % Volatile | 75 | 70 | 75 | 80 | 65 | 81 | 90 | 80 |
| Waste Solids (lb/million gal.) | 161 | 40 | 144 | 250 | 680 | 230 | 193 | 253 |
| Volatile (lb/million gal.) | 121 | 28 | 108 | 200 | 441 | 202 | 174 | 202 |
| Effluent Solids | | | | | | | | |
| (lb/million gal.) | 296 | 445 | 166 | 204 | 290 | 470 | 197 | 483 |
| Volatile (lb/million gal.) | 198 | 245 | 113 | 141 | 189 | 342 | 118 | 400 |

\+ Peripheral feed - no center feed section
Area at surface 96 $ft^2$ X 6 ft. deep
Area below feed skirt 107 $ft^2$ X 5 ft. deep
Total depth 11 ft. deep

± Center feed section area at surface 78 $ft.^2$ X 4 ft.deep
area above feed skirt 96 $ft^2$ X 2 ft.deep
area below feed skirt 107 $ft^2$ X 5 ft.deep

TABLE 4 (CONTINUED)

CLARIFIER VARIABLES AND OPERATING CONDITIONS
BLUE PLAINS

| 9 March 1-18 | 10 April | 11 May | 12 June | 13 July | 14 August | 15 September | 16 October | 17 November | 18 December 1-21 | 19 December 22-31 |
|---|---|---|---|---|---|---|---|---|---|---|
| 950 ± | - @ | - @ | - @ | - @ | - @ | - @ | - @ | - @ | - @ | - @ |
| 780 | (290-620) | 560 | (280-650) | 975 | 975 | 975 | 975 | 975 | 975 | 975 |
| 700 | - | - | - | - | - | - | - | - | - | - |
| 950 | - | - | - | - | - | - | - | - | - | - |
| 780 | (290-620) | 560 | (280-650) | 650 | 650 | 650 | 650 | 650 | 650 | 650 |
| 700 | - | - | - | - | - | - | - | - | - | - |
| 35 | 17 | 21 | 24 | 54 | 61 | 58 | 47 | 36 | 33 | 32 |
| 81 | 120-190 | 265 | 173 | 50 | 30-35 | 33 | 42 | 36 | 40 | 56 |
| 0.79 | 0.85 | 0.78 | 1.28 | 1.92 | 2.22 | 2.38 | 2.26 | 2.05 | 2.41 | 2.30 |
| 77 | 80 | 77 | 73 | 70 | 70 | 70 | 71 | 79 | 81 | 82 |
| 130 | - | 720 | 142 | 168 | 253 | 460 | 340 | 178 | 200 | 310 |
| 100 | - | 550 | 104 | 118 | 178 | 323 | 242 | 140 | 160 | 254 |
| 410 | - | 100 | 88 | 92 | 133 | 160 | 130 | 194 | 190 | 140 |
| 375 | - | - | - | - | 100 | 104 | 99 | 130 | 123 | 88 |

@ Two center feed clarifiers at 78 $ft^2$ each X 11 feet deep
feed sections at 3 $ft^2$ each X 3.5 ft deep

As expected with filamentous growth in late Spring 1971, allowable clarifier overflow rates were markedly decreased as shown in Table 4. Again, during the current 1971-72 Winter, maximum overflow rates of 975 gpd/ft$^2$ have been demonstrated at Blue Plains.

SOLIDS HANDLING

The other integral part of the oxygen system is the excess solids handling equipment. Of utmost importance is the relative ease with which the oxygen activated sludge process can be operated in endogenous respiration, thereby substantially reducing the quantity of excess sludge to be handled. This factor will reduce the number and/or size of the selected sludge handling and disposal facilities. However, the increased operating costs resulting from the increased oxygen necessary to oxidize ("burn-up") the excess sludge and the larger reactor/clarifier capabilities needed to hold the increased solids inventory required for endogenous respiration must be balanced economically with the reduction in size of the solids handling and disposal units.

Another factor to be considered is that the larger the clarifier volume, the thicker the underflow solids concentration. It may be economically feasible to properly size the reactor/clarifier combination to yield underflow solids sufficiently thick to be dewatered directly without prior thickening or digestion. This would be accomplished by selecting a small reactor and large clarifier. An alternative to the above is to select a large reactor and a small clarifier and provide additional thickener capabilities. In the EPA/DC Pilot Plant, the excess solids are thickened separately by air flotation or gravity thickening. These solids have been thickened to over 4.5% without chemical additives by both gravity and air flotation thickening.

CONCLUSIONS

1. A gas-tight biological oxygen reactor with independent control of dissolved oxygen and mixing, coupled with an aerobic clarifier, produces a good quality secondary effluent on District of Columbia primary effluent with 1.5 to 2.5 hours average detention time (based on raw flow) with MLSS concentrations between 4000 and 8000 mg/l.

2. Biodegradable organics in the Blue Plains primary effluent are essentially completely insolubilized by the oxygen process (less than 5 mg/l of soluble BOD). Total carbonaceous BOD removal depends upon the amount of suspended solids in the effluent and, therefore, on the ability to clarify.

3. Oxygen micro-organisms are visually the same as those in a typical conventional system; however, the rate of activity of the oxygen volatile solids is greater above an SRT of 6 days.

4. Oxygen activated sludge is subject to filamentous (Sphaerotilis) growth as similarly observed in the air systems when operated below an SRT of 5 days on DC primary effluent.

5. Sludge in the oxygen system underflow settles to approximately 1.0-1.4% solids in a clarifier with 1.9 hours of hydraulic retention time and 2.0% to 2.4% in a clarifier with 2.8 hours of hydraulic retention time.

6. Total production of excess biological solids is significantly lower in the Blue Plains oxygen system than in a parallel step aeration system at SRT's above 6 days with as little as 0.35 lb of excess solids produced/lb BOD added at an SRT of 13 days.

7. When the oxygen clarifier is operated with a deep feed well or with the mixed liquor sufficiently concentrated to settle in a subsidence (zone) settling pattern, the blanket acted as a filter and produced high quality effluent.

In the Summer and Fall, 1970, the clarifier operated at a peak rate of 1940 gpd/ft$^2$. In the 1970-71 Winter, oxygen clarifier rates could not exceed a sustained 975 gpd/ft$^2$. A larger clarification area is required in the Winter than in the Summer on District of Columbia wastewater for a given MLSS concentration.

8. With a shallow center feed well and with the mixed liquor concentrations low enough (under 4500 mg/l) to permit discrete particle settling, better settling rates are observed in the oxygen clarifier than with the method of clarifier operation described in No. 7 above. Only moderate decreases in effluent quality (increase in SS from 15 to 25 mg/l) are observed with this type of clarifier operation at Blue Plains.

9. Nitrification is achieved in the oxygen aeration system in the Summer and Fall at Blue Plains.

10. Average effluent phosphorus residuals of 1.8 mg/l as P with an alum dosage of 1.4 $Al^{+++}$ to 1.0 P, by weight, were achieved in the oxygen system. Higher phosphorus removals are possible with higher alum dosages, but in areas with moderate wastewater alkalinity, pH control may be required to prevent the depletion of alkalinity reserves in the oxygen system.

11. Based upon the influent and exhaust gas flows, over 95% of the input oxygen is consistently utilized in the Blue Plains oxygen reactor.

SUMMARY

Again, the oxygen activated sludge system must be viewed as an entirely unique approach, and compared on a total system basis with other alternative systems. Reactor and clarifier sizing must be coordinated. As the reactor size is increased, a lower MLSS concentration is required for a given biological state (F/M ratio). The lower the MLSS concentration, the smaller the required secondary clarification area. The solids handling requirement of an oxygen system depend

on the biology established in the reactor/clarifier combination. Thus, if minimum excess biological sludge production is required, then more capacity is required in the reactor/clarifier combination. Further, the concentration of the clarifier underflow solids is dependent on clarifier volume.

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