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# Project Summary

# Treatment of Municipal Wastewaters by the Fluidized Bed Bioreactor Process

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**A 2-year, large-scale pilot investigation was conducted at the City of Newburgh Water Pollution Control Plant, Newburgh, NY, to demonstrate the application of the fluidized bed bioreactor process to the treatment of municipal wastewaters. The reactor was 1.2 m wide, 1.8 m long, and 6 m high, with a typical active bed volume of 9 $m^3$. The bed medium was quartzite sand with a nominal diameter of 0.5 mm. Oxygenation was accomplished by a device installed below grade (providing approximately 14 m static head); the oxygen source was on-site liquid oxygen tanks. The wastewater feed was primary effluent drawn from the main plant.**

**The experimental effort investigated the ability of the process to treat municipal wastewater to secondary levels. Process design elements such as removal rates, temperature effects, oxygen transfer rates, oxygen requirements, sludge production, and bed maintenance and control were evaluated. Additionally, the studies evaluated the stability of the process under high hydraulic peak loads typical of combined sewer systems.**

**The plant primary effluent had an average 5-day biochemical oxygen demand (BOD) of 103 mg/L, a chemical oxygen demand (COD) of 287 mg/L, and an average total suspended solids (TSS) of 109 mg/L. An organic loading rate of 0.2 to 0.3 gm BOD/gm volatile suspended solids/day is suggested in order to meet secondary treatment levels. Suspended solids removals are relatively low through the reactor; the results indicate that final clarification would be required for the plant to meet a 30 mg/L TSS limitation. The oxygenator exhibited good performance (greater than 80% oxygen transfer) at gas to liquid ratios less than 0.04.**

**The system exhibited good stability under high short-term peak hydraulic loadings. Maximum to average raw flow ratios up to 5 could be handled without bed washout. Special reactor modifications to yield an expanded upper section allowed a ratio as high as 10 to 1 without loss of the bed. The system exhibited relatively quick recovery to normal treatment efficiency after the hydraulic surges.**

**Applications appropriate to the process are for secondary treatment and as a roughing treatment process in plant upgrades.**

***This Project Summary was developed by EPA's Water Engineering Research Laboratory, Cincinnati, OH, to announce key findings of the research project that is fully documented in a separate report of the same title (see Project Report ordering information at back).***

## Introduction

Fluidized biological reactors represent a newly developed and innovative technology that combines the advantages of activated sludge and trickling filtration. It attempts to combine the efficiency of the suspended growth system with the stability and efficient operation of the attached growth process. Within the fluidized bed reactor, a fixed film of biologic growth is allowed to develop on the surface of a granular medium, such as sand. Wastewater, which has been oxygenated, flows upward through the medium at velocities sufficient to expand the bed. This expansion, or fluidization, occurs when the drag due to the upward fluid velocity is greater than the downward force due to gravity. The fluidization separates the particles and allows microorganisms to attach and grow on all surfaces of the medium. Thus, the very large surface area available can produce high volatile solids concentrations, often several times greater than in suspended growth reactors. The attachment of microorganisms to the medium also reduces subsequent requirements for wastewater clarification.

A portion of the effluent stream is recycled; the ratio of recycle to raw flow affects both column expansion and oxygen transfer, and often serves as a process control parameter. In the majority of reported studies, the reactors had been operated on constant flow regimes, with recycle ratios ranging from zero to three for municipal sewage. The recycle mode can also provide stability in applications with high influent hydraulic variability.

Control of the biomass film thickness and sludge wasting are accomplished by a sand-biomass separation device. This generally operates by shearing the biomass from the sand, separating the sludge from the sand, then returning the sand to the reactor.

One advantage of the fluidized bed process is its reduced land area requirement, due to its ability to carry a high biomass inventory. Typical volumetric organic loading rates of 3 to 20 kg BOD/$m^3$-d and mass organic loading rates of 0.2 to 2.0 kg BOD/kg VSS-d (where VSS is volatile suspended solids) have been reported. Typical rates for conventional activated sludge treatment are 0.3 to 2 kg BOD/$m^3$-d and 0.2 to 0.6 kg BOD/kg VSS-d, respectively.

In this project a large-scale pilot study was conducted at the Newburgh, NY, Water Pollution Control Plant to investigate the application of the fluidized bed bioreactor process to the treatment of municipal wastewaters. The study was directed to the system's ability to achieve secondary treatment levels of 30 mg/L TSS and BOD, and its stability under the hydraulic variability associated with a plant receiving combined sanitary and storm wastewaters. A large-scale Dorr-Oliver Oxitron* pilot plant was set up on-site and continuously monitored for treatment performance; specific studies and tasks were also directed to assessing the unit operations that comprise the process.

## Program Objectives and Scope of Work

The study included the installation of a fluidized bed bioreactor pilot system. This system was operated over a period of approximately 18 months, receiving settled primary effluent from the Newburgh Water Pollution Control Plant. Sampling of influent, effluent, and the reactor was conducted to assess performance under variable loading conditions and to evaluate bed stability. Specific task elements addressed bed maintenance requirements, solids settleability, oxygenation capacity, and bed wasting and sand-biomass separation procedures.

The objectives of the program are as follows:

1. To demonstrate the ability of the fluidized bed bioreactor process to achieve secondary treatment levels (30 mg/L TSS and BOD);
2. To assess the stability of the process and its ability to handle the highly variable hydraulic loads characteristic of a combined sanitary-stormwater collection system during rainfall runoff events;
3. To evaluate the operation and maintenance requirements of the fluidized bed process as applied to the treatment of a municipal treatment plant primary effluent;
4. To define and establish the major process parameters that influence the design and operation of the fluidized bed reactor, such as hydraulic and organic loading rates, oxygen transfer and utilization efficiencies, biomass-sand separation efficiency, bed control, sludge production, and wasting requirements; and
5. To develop cost information for the fluidized bed process relative to its application to municipal wastewaters.

## Description of Facilities

The fluidized bed pilot plant facility was located at the Newburgh Water Pollution Control Plant, Newburgh, NY. The city is approximately 100 kilometers (60 miles) north of New York City. The treatment plant is a conventional secondary activated sludge facility, treating domestic and industrial wastes, and discharging to the Hudson River. The plant services a population of approximately 20,000; industrial inputs are primarily dye mills and commercial laundry wastewaters.

## *Description of the Pilot Plant*

The Oxitron pilot plant was a skid mounted, field assembled plant designed by Dorr Oliver, Inc., Stamford, CT, to demonstrate the capabilities of fluidized bed reactors. This same system had previously been used to demonstrate nitrification of domestic wastewaters at the Greenwich Wastewater Treatment Plant in Greenwich, CT. The facility was situated in an area of the Newburgh Water Pollution Control Plant that had been set aside for future expansion.

Figure 1 presents a process schematic of the pilot plant showing all major unit operations. The 1,900-L influent head tank located near the primary effluent channel was consistently fed by one of two submersible pumps. Primary effluent was pumped to the tank at a rate generally in excess of 1,500 L/min. Excess flow passed through an overflow pipe back into the primary effluent channel. A float valve was provided to activate a low level alarm and automatically shut off the pilot plant system if the level of the head tank dropped as a result of interrupted feed supply. A 25-cm-diameter PVC pipe supplied feed from the head tank to the influent pump located in the pilot plant building. The influent pump was an Aurora, 23-cm, 15-hp centrifugal pump with a pressure relief bypass loop. Normal operating pressure of this pump was 2.8 to 3.2 kg/$cm^2$. Maximum flow was approximately 950 L/min, depending on system pressure.

The raw wastewater feed was then passed through a rotary strainer to remove large solids and protect the reactor distributor nozzles from plugging. The strainer was backwashed twice per day or more frequently as needed. Pressure gauges on both sides of the strainer were used to monitor its performance. Generally, under normal operation conditions, a pressure drop of under 0.2 kg/$cm^2$ was maintained.

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*Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

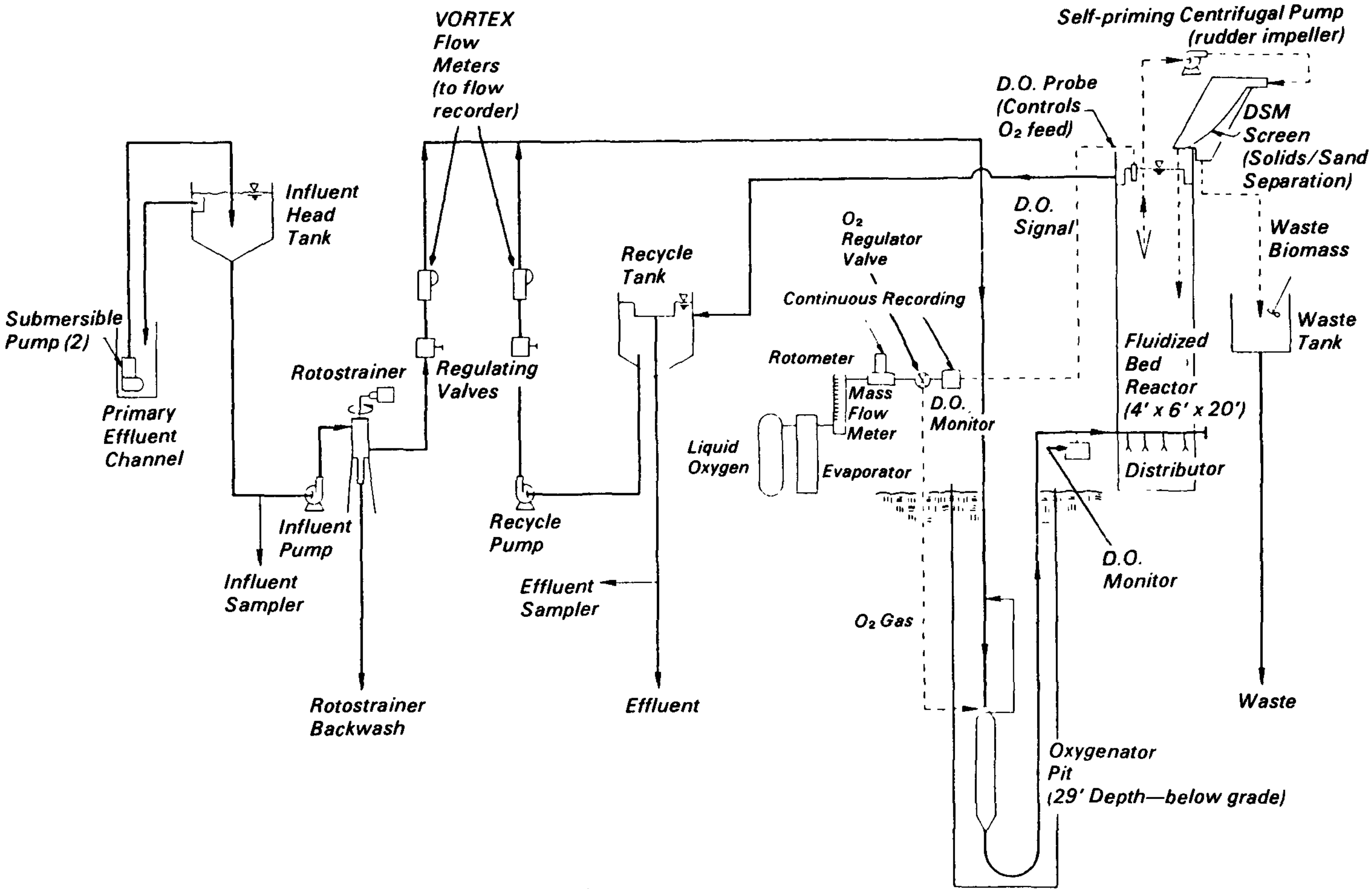


*Figure 1. Process schematic of the fluidized bed pilot facility.*

Feed then passed through a Keystone control valve and Ball Vortex flow meter. The control valve could be manually or automatically operated. The automatic mode was used to provide diurnal flow variation similar to that of the Newburgh treatment plant flow. A signal from the Newburgh plant flow meter was used to drive the pilot plant control valves. The flow meter was a circulating cage type that proved to be very reliable. Occasionally, the flow meters were electronically calibrated and balanced. The meters were also connected to remote flow recorders and totalizers at the control panel.

The raw wastewater, combined with the recycle waters, then flowed to the oxygenator where oxygen gas was injected to satisfy oxygen requirements for aerobic biological treatment. The oxygenator provided gas/liquid contact to maximize dissolution of oxygen into the influent wastewater. At the inlet to the oxygenator, the combined influent and recycle streams were mixed under pressure with gaseous oxygen. The oxygenator was below grade to provide approximately 14 m of static head (the difference between the top of the reactor and the bottom of the oxygenator pit). This pressure enhanced oxygen transfer and allowed dissolved oxygen levels of up to 60 mg/L to be achieved. The internal design of the oxygenator ensures that gas/liquid contact is maximized for efficient oxygen transfer. The gas recycle loop allows recirculation of undissolved oxygen.

The biological support medium was quartzite sand with a median particle size of approximately 0.50 mm. At start-up, the reactor was charged with 4 m$^3$ of medium. Sand was also added to the reactor throughout the study to compensate for attrition losses.

The reactor was a 1.2 m by 1.8 m steel tank, 6.1 m high. The tank was comprised of four flanged sections, the uppermost and lower two sections had clear acrylic windows on one side to allow visual observation. A scale was added to the windows to indicate bed height. The zero point of the scale was 32 cm from the bottom to compensate for the distributor piping and plates. The distributor plates directed the flow upward; note that the nozzles were pointed downward. Two 1.8-m-long, multi-V-notch weir plates were located 5.5 m from the bottom of the tank. Effluent dissolved oxygen was measured by a Beckman dissolved oxygen analyzer.

A portion of the pilot plant effluent was recycled to the reactor to reduce the strength of the incoming wastewater when necessary, and to maintain bed fluidization during low flow periods and during influent flow interruptions. The recycle tank split the reactor overflow into a recycle stream and an effluent stream. The reactor overflow was introduced into the recycle tank through a tangential inlet to set up a circulating pattern within the tank. This action imparted an outward velocity to any solid particles entering the recycle tank.

Centrifugal forces push the sand particles, in the event that they should overflow the reactor, to the periphery of the tank where they settle out into the conical section of the tank. A bottom drain was provided to allow for cleaning this accumulation from the tank.

Effluent left the recycle tank by overflowing the center launder. Recycle flow was withdrawn from the tank through a submerged outlet in the conical tank section, and was pumped through a Keystone butterfly flow control valve, a Ball Vortex flow meter, and combined with the influent raw wastewater before entering the oxygenator. A check valve was located in the recycle line to prevent backflow during periods when the recycle pump was shut down.

A ratio control station was installed in the pilot plant to allow pilot plant flows to vary as the Newburgh wastewater treatment plant flow varied. The operator had the option of operating the pilot plant under constant influent and recycle flow or an automatic mode where a signal from the Newburgh flow meter controlled the pilot plant flow. The automatic mode could be operated in two ways: constant recycle with varying influent flow, providing a variable hydraulic flux; or constant total flow, with recycle and influent flow varying to provide a constant hydraulic flux.

The bed wasting operation is critical for effective bed control and maintenance of thin biofilms. A rubber-lined centrifugal pump transported the sand and biomass to a gravity screening device (Dorr-Oliver DSM) and provided the shearing force necessary to dislodge the biomass from the sand media. With the aid of spray water, sheared biomass passed through the screen and sand media was retained and returned to the reactor. The waste biomass was collected in the waste tank and pumped to the main plant's recycle pit after volume measurement and sampling for TSS and VSS.

### *Small-Scale Pilot Plant*

As previously stated, the maximum feed flow from the reactor influent pump was approximately 950 L/min, depending on the operating pressure of the reactor. The lowest available flow (to maintain fluidization) was approximately 380 L/min (176 L/min per $m^2$ based on influent flow). This provides a maximum to average flow ratio of 2.5:1. One goal of the project was to test the system under higher maximum to average ratios than was possible with the large pilot system. Dorr-Oliver supplied a smaller pilot plant to evaluate stormwater impact. This smaller unit is shown on Figure 2. The reactor consisted of two sections of clear acrylic pipe. The lower section was 15 cm in diameter, while the upper section was interchangeable. Two upper sections were used, one 15-cm-diameter section to provide a straight column, and one 23-cm section to provide an expanded column.

Influent to the column was taken from the pilot plant distributor and went through a booster pump and rotometer. This configuration permitted a maximum hydraulic loading of 2,100 L/min per $m^2$, corresponding to a ratio of 12:1 when using the pilot plant minimum influent hydraulic flux for fluidization.

A small DSM screen was set up to accomplish wasting from the small unit. Wasting was done manually.

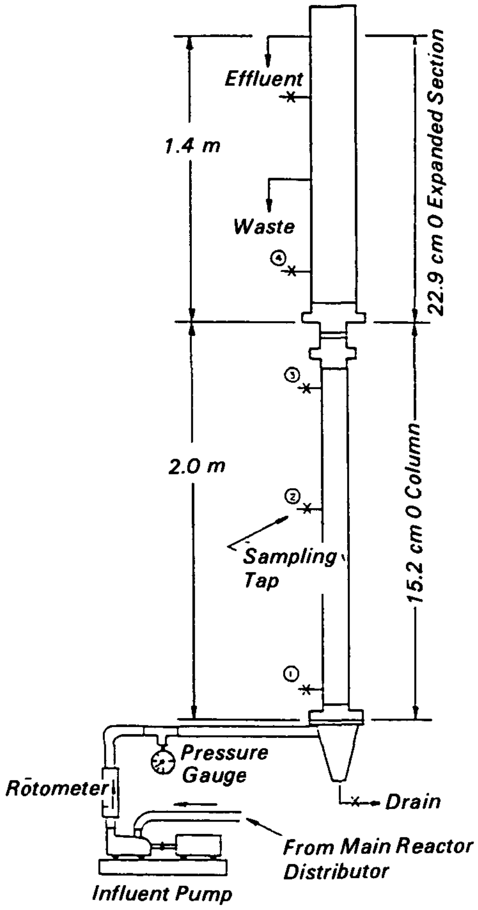


***Figure 2.*** *Schematic of small pilot unit used during a portion of the studies.*

## Experimental Program

The experimental phase of the project covered a total period of 19 months, from September 1983 through March 1985. The reactor was operated under constant raw wastewater conditions from September 1983 through September 1984. During the final portion of the study, December 1984 through March 1985, the raw flow was automatically adjusted to be proportional to the Newburgh plant's raw flow rate. The recycle, in this case, was automatically adjusted in order to maintain a constant flux through the reactor. Operating periods and number of samples taken were as follows:

1. 09/15/83 to 11/10/83, 28 samplings;
2. 02/06/84 to 03/08/84, 31 samplings;
3. 03/21/84 to 04/09/84, 15 samplings;
4. 04/16/84 to 05/22/84, 28 samplings;
5. 06/26/84 to 07/24/84, 22 samplings;
6. 08/16/84 to 08/21/84, 4 samplings;
7. 08/29/84 to 10/09/84, 31 samplings;
8. 12/19/84 to 01/04/85, 8 samplings;
9. 01/21/85 to 02/08/85, 10 samplings;
10. 02/14/85 to 02/20/85, 4 samplings; and
11. 02/28/85 to 03/14/85, 7 samplings.

## Experimental Results

Average results for the 11 periods of operation are shown in Table 1. Retention times, based on bed height, but assuming empty bed volume, ranged between 20 and 40 min. High volatile solids inventories and use of pure oxygen permit these short contact times. BOD removals, based on total influent BOD, but soluble effluent BOD, were above 70% with the exception of one time period. BOD removals were based on

**Table 1.** *Reactor Performance Summary Table§*

| Period Number | 1 | 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| Dates | 9/15/83-<br>11/10/83 | 2/6/84-<br>3/8/84 | 3/21/84-<br>4/9/84 | 4/16/84-<br>5/22/84 | 6/26/84-<br>7/24/84 | 8/16/84-<br>8/21/84 |
| Number of Weeks | 9 | 5 | 4 | 6 | 5 | 1 |
| <u>Hydraulic Data</u> | | | | | | |
| Raw Flow (L/min) | 335 | 305 | 332 | 341 | 311 | 417 |
| Recycle Flow (L/min) | 664 | 580 | 641 | 488 | 625 | 647 |
| Total Flow (L/min) | 998 | 885 | 962 | 834 | 936 | 1064 |
| Ratio Recycle/Raw | 2.0 | 1.9 | 1.9 | 1.4 | 2.0 | 1.6 |
| Hydraulic Flux (m/min) | 0.46 | 0.41 | 0.44 | 0.39 | 0.43 | 0.49 |
| Retention Time* - RawFlow (min) | 25.1 | 29.0 | 29.9 | 27.9 | 27.1 | 22.3 |
| <u>Biochemical Oxygen Demand-5 Day (BOD)</u> | | | | | | |
| BOD Influent (mg/L) | 138.4 | 71.7 | 37.8 | 56.1 | 71.9 | 125.5 |
| SBOD Influent (mg/L)** | 59.5 | 44.0 | 23.3 | 28.5 | 33.2 | 34.3 |
| BOD Effluent (mg/L) | 87.6 | 38.5 | 17.3 | 44.5 | 42.6 | 45.8 |
| SBOD Effluent (mg/L) | 32.2 | 19.6 | 10.4 | 12.8 | 15.9 | 8.5 |
| BOD Loading Rate (kg/d/kd VS) | 0.44 | 0.26 | 0.11 | 0.20 | 0.36 | 0.63 |
| BOD Loading Rate (kg/d/$m^3$) | 7.72 | 3.42 | 1.31 | 2.56 | 3.55 | 7.53 |
| SBOD Loading Rate (kg/d/kg VS) | 0.19 | 0.16 | 0.07 | 0.11 | 0.16 | 0.18 |
| SBOD Loading Rate (kg/d/$m^3$ | 3.4 | 2.1 | 0.83 | 1.41 | 1.6 | 2.1 |
| BOD Removal (Percent)† | 77.2 | 71.1 | 60.4 | 72.7 | 74.8 | 87.4 |
| BOD Removal Rate (kg/d/kg VS)† | 0.37 | 0.21 | 0.09 | 0.16 | 0.29 | 0.14 |
| BOD Removal Rate (kg/d/$m^3$) | 6.05 | 2.81 | 1.01 | 1.86 | 2.95 | 2.63 |
| SBOD Removal Rate (kg/d/kg VS)# | 0.11 | 0.11 | 0.04 | 0.06 | 0.10 | 0.05 |
| SBOD Removal Rate (kg/d/$m^3$) | 1.46 | 1.37 | 0.44 | 0.70 | 0.83 | 1.60 |
| <u>Suspended Solids (SS)</u> | | | | | | |
| Total Suspended Solids (mg/L) | | | | | | |
| - Influent TSS | 116.4 | 65.8 | 50.5 | 64.4 | 75.9 | 44.5 |
| - Effluent TSS | 71.1 | 44.4 | 29.5 | 47.8 | 33.2 | 38 |
| Volatile Suspended Solids (mg/L) | | | | | | |
| - Influent VSS | 87.3 | 44.3 | 33.3 | 50.5 | 54.9 | 33.5 |
| - Effluent VSS | 56.6 | 31.2 | 22.1 | 37.5 | 24.1 | 30.0 |
| TSS Removal (percent) | 39.0 | 32.5 | 41.5 | 25.8 | 56.3 | 15.0 |
| <u>Bed Characteristics</u> | | | | | | |
| Total Solids (kg) | 5750.0 | 4540.0 | 4332.0 | 4662.0 | 4060.0 | 5034.0 |
| Total Solids (gm/L) | 565.0 | 493.0 | 420.0 | 475.0 | 420.0 | 554.0 |
| Total Volatile Solids (kg) | 149.0 | 123.0 | 119.0 | 128.0 | 92.00 | 115.0 |
| Total Volatile Solids (gm/L) | 16.8 | 13.3 | 11.5 | 12.8 | 9.6 | 12.6 |
| Calculated Film Thickness (µm) | 28.3 | 123.0 | 78.0 | 79.5 | 145.0 | 15.0 |
| Total Bed Height (m) | 3.9 | 4.1 | 4.6 | 4.4 | 3.9 | 4.3 |
| Percent Fluidization | 157.0 | 240.0 | 298. | 262.0 | 299.0 | 200.0 |

§Note that calculated values are accomplished on single points, and then averaged.

*Based on empty bed volume, at average bed height.

**SBOD is soluble BOD; SCOD is soluble COD.

†BOD or COD removal determined by total influent minus soluble effluent.

#SBOD or SCOD removal determined by soluble influent minus soluble effluent.

*Table 1.* *(Continued)*

| Period Number | 7 | 8 | 9 | 10 | 11 |
|---|---|---|---|---|---|
| Dates | 8/29/84-<br>10/9/84 | 12/19/84-<br>1/4/85 | 1/21/85-<br>2/8/85 | 2/14/85-<br>2/20/85 | 2/28/85-<br>3/14/85 |
| Number of Weeks | 7 | 3 | 3 | 1 | 3 |
| Hydraulic Data | | | | | |
| Raw Flow (L/min) | 382 | 413 | 375 | 384 | 466 |
| Recycle Flow (L/min) | 622 | 636 | 665 | 511 | 448 |
| Total Flow (L/min) | 1005 | 978 | 913 | 915 | 914 |
| Ratio Recycle/Raw | 1.6 | 1.5 | 1.8 | 1.3 | 1.0 |
| Hydraulic Flux (m/min) | 0.46 | 0.45 | 0.42 | 0.42 | 0.42 |
| Retention Time* - RawFlow (min) | 24.3 | 24.1 | 27.1 | 23.6 | 19.9 |
| Biochemical Oxygen Demand-5 Day (BOD) | | | | | |
| BOD Influent (mg/L) | 121.1 | 128.6 | 151.9 | 111.5 | 120.6 |
| SBOD Influent (mg/L)** | 54.3 | 62.1 | 72.5 | 59.5 | 59.1 |
| BOD Effluent (mg/L) | 66.5 | 95.0 | 83.7 | 55.0 | 50.6 |
| SBOD Effluent (mg/L) | 17.1 | 33.5 | 26.2 | 22.0 | 24.0 |
| BOD Loading Rate (kg/d/kd VS) | 0.49 | 0.89 | 0.56 | 0.59 | 0.54 |
| BOD Loading Rate (kg/d/$m^3$) | 6.86 | 8.00 | 7.80 | 7.74 | 8.28 |
| SBOD Loading Rate (kg/d/kg VS) | 0.20 | 0.44 | 0.27 | 0.33 | 0.26 |
| SBOD Loading Rate (kg/d/$m^3$ | 2.9 | 3.9 | 3.8 | 4.3 | 4.1 |
| BOD Removal (Percent)† | 86.9 | 73.9 | 82.6 | 79.3 | 80.5 |
| BOD Removal Rate (kg/d/kg VS)† | 0.36 | 0.32 | 0.43 | 0.27 | 0.49 |
| BOD Removal Rate (kg/d/$m^3$) | 5.27 | 2.95 | 5.98 | 5.65 | 6.55 |
| SBOD Removal Rate (kg/d/kg VS)# | 0.13 | 0.11 | 0.17 | 0.09 | 0.15 |
| SBOD Removal Rate (kg/d/$m^3$) | 1.84 | 1.09 | 2.24 | 0.41 | 2.38 |
| Suspended Solids (SS) | | | | | |
| Total Suspended Solids (mg/L) | | | | | |
| - Influent TSS | 138.6 | 126.3 | 154.8 | 116.8 | 94.6 |
| - Effluent TSS | 65.3 | 88.8 | 97.0 | 80.3 | 62.0 |
| Volatile Suspended Solids (mg/L) | | | | | |
| - Influent VSS | 98.2 | 95.8 | 122.1 | 83.8 | 73.0 |
| - Effluent VSS | 47.6 | 65.7 | 78.0 | 69.8 | 47.7 |
| TSS Removal (percent) | 53.0 | 29.7 | 37.3 | 31.3 | 34.0 |
| Bed Characteristics | | | | | |
| Total Solids (kg) | 5238.0 | 5865.0 | 4938.0 | 4919.0 | 5350.0 |
| Total Solids (gm/L) | 548.0 | 543.0 | 449.0 | 534.0 | 523.0 |
| Total Volatile Solids (kg) | 136.0 | 97.0 | 150.0 | 114.0 | 158.0 |
| Total Volatile Solids (gm/L) | 14.2 | 8.9 | 13.6 | 12.5 | 15.9 |
| Calculated Film Thickness (μm) | 14.5 | 88.8 | 98.9 | 66.8 | 118.4 |
| Total Bed Height (m) | 4.3 | 4.6 | 4.7 | 4.2 | 4.3 |
| Percent Fluidization | 205.0 | 200.0 | 279.0 | 212.0 | 221.0 |

§Note that calculated values are accomplished on single points, and then averaged.
*Based on empty bed volume, at average bed height.
**SBOD is soluble BOD; SCOD is soluble COD.
†BOD or COD removal determined by total influent minus soluble effluent.
#SBOD or SCOD removal determined by soluble influent minus soluble effluent.

soluble effluent BOD because the existing system did not provide adequate suspended solids removal. Final clarification is considered necessary to assure effluent suspended solids below 30 mg/L. The unusually high BOD removals experienced in periods 6 and 7 occurred after the reactor was recharged with clean sand and are attributed to development of a dense, thin biofilm.

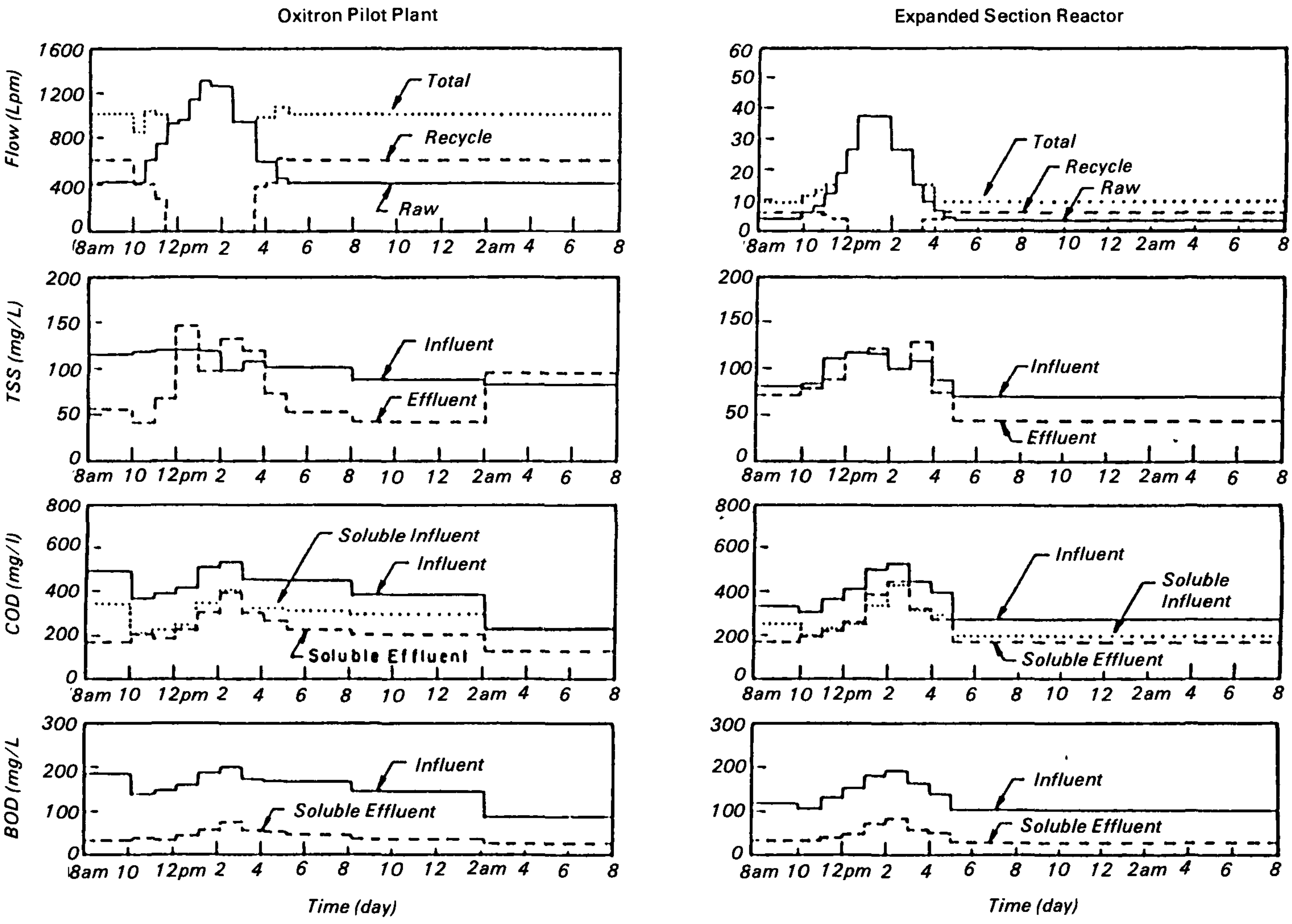
A number of short-term runs were made using the small pilot unit with expanded top sections shown in Figure 2. Results of a run made on February 20, 1985, are shown in Figure 3. The maximum hydraulic flux attained in the lower section of the column was over 2,000 L/min per m² with the bed being maintained about 0.5 m below the point of washout. Although the degree of pollutant removal was small during high flux, the bed could be contained in the reactor and the biofilm was not lost from the sand particles. The large pilot plant was run on February 20 at the highest feed rate possible with available pumping capacity, but the maximum flux was only 500 L/min per m². Pollutant removal was better from this system than from the small pilot system.

## Conclusions

The fluidized bed aerobic bioreactor process appears technically viable for application to the secondary treatment of municipal wastewaters. Improvements must be made to the sand-biomass separation system for more effective operation. Estimated installed and operating costs are in the range of other biological treatment methods. Organic loadings between 0.2 and 0.3 kg BOD/kg VSS-day would be required to yield soluble effluent BOD levels less than 20 mg/L. Final clarification will be necessary to assure effluent suspended solids levels less than 30 mg/L.

The process is particularly suited for applications in which land is limited, and where roughing treatment is appropriate. In this case, the process, at loadings of 0.4 to 0.5 kg BOD/kg VSS-d can be expected to accomplish greater than 50% BOD removal. Intermediate clarification would not be needed if followed by a conventional biological system for final treatment.

There is a correlation of effluent suspended solids concentration with the TSS and total BOD loading to the reactor. Impractically low loadings would be required in order to meet secondary effluent TSS limits without final clarifi-


Oxitron Pilot Plant

Expanded Section Reactor

*Figure 3.* *Flow and influent/effluent COD, BOD, and TSS data from the February 20, 1985 hydraulic variability evaluation.*

cation. Although the effluent TSS levels were adversely affected at Newburgh by the inefficiencies in the bed wasting system, final clarification should still be required. High rate clarification (50 $m^3/m^2/d$) is feasible if coupled with chemical conditioning. Doses of 10 mg/L alum and 1 mg/L polymer were found to be effective for the Newburgh application.

The process has the ability to undergo significant short term hydraulic peak loads and still maintain bed integrity. Removal effectiveness is, however, greatly reduced. A straight wall reactor can operate at a hydraulic flux of 1 m/min without any loss of bed, approximately 2.5 times its normal operating flux. This is equivalent to a maximum to average raw flow ratio of 4 to 6, effected by substituting raw flow for recycle flow.

Modification of the reactor to one with an expanded upper section (approximately 2.3 times the cross-sectional area of the lower section) will significantly enhance the reactor's ability to undergo hydraulic surges. A flux of 2.5 m/min was demonstrated (based on lower section cross-sectional area), with no loss of bed. This is equivalent to a maximum to average raw flow ratio of 10. The reactor organic removal efficiencies return to normal relatively quickly after cessation of the hydraulic surge.

Bed stability can be controlled by the hydraulic flux and the average particle density. The particle density will be a function of the average biofilm thickness; this should be maintained below 50 $\mu$m. Effective control requires maintenance of adequate sand and volatile solids levels; the bed must be monitored by direct sampling and analysis to accomplish this.

The sand concentration should average 400 to 600 g/L through the bed, under expanded conditions, with an average volatile solids concentration of 10,000 to 15,000 mg/L. The flux should be controlled to maintain a bed expansion of 50% to 100%.

Maximum oxygen transfer efficiencies can be accomplished with the oxygenator at gas to liquid ratios less than 0.03 L/min $O_2$ at standard conditions per L/min wastewater.

A key element to the successful operation of the fluidized bed is the ability to maintain and control optimum reactor bed conditions. The feed to the reactor must be kept free of debris; this is accomplished by screening the primary effluent (or combined primary effluent/reactor recycle). The distributor must give even distribution of flow across the cross-sectional area of the reactor, without localized velocity gradients. The distributor ports must be non-clogging; pressure drops should be monitored across the distributor and the system should provide a means to access the distributor ports for cleaning.

The bed wasting operation is critical for effective bed control and maintenance of thin biofilms. Centrifugal pumps with rubber-lined impellers provide good shearing of the biomass from the sand and the gravity screening device (Dorr-Oliver DSM) used in this study provided good separation. Spray water is essential for effective operation of the separation screen.

The bed wasting operation was labor intensive and subject to significant equipment problems. The problems related primarily to pumping the sand/biomass mixture without clogging, and materials damage. The operation can also affect the effluent suspended solids levels from the reactor because of inefficiencies in separating the biomass from the sand before the sand is returned to the reactor. Improvements to the existing system would include a submersible pump to remove the sand/biomass (and shear the biomass from the sand), a spray water system to enhance removal of the biomass by the separation screen, and baffles on the screen to prevent flow channelling.

An alternative method would be to install a system separate from the reactor. An agitator would shear the biomass, and a two-stage rotational separator and gravity settler would provide for separation of the sand from the biomass and removal of the biomass and sand slurries.

## Recommendations

The applications appropriate to the fluidized bed process are as a roughing treatment system prior to a conventional treatment system, or as a secondary treatment system for carbonaceous BOD removal. It is recommended that the process be given particular consideration when there is limited land available, significant hydraulic variability, and/or a proximate, cost-effective source of pure oxygen. Clarification should be provided if the process must meet secondary effluent TSS limits.

In cases where high hydraulic variability is expected, it is recommended that the modified reactor with the expanded top section be used. This can increase the capacity of the system and significantly enhance bed stability during short term hydraulic peak loads.

Further work and subsequent demonstration is recommended with regard to bed wasting and control of bed characteristics. This should address procedures and equipment for removing bed sand/biomass, shearing the excess biomass from the sand, separation of the sand and biomass, and effective pumping/transport of the sand/biomass and sand slurries. The efforts should be directed to reducing the labor requirement and to designing more effective equipment for accomplishing these tasks. Consideration should be given to wholly separating this process from the reactor.

Additional work is needed to demonstrate the performance of the bioreactor during the transient conditions of peak hydraulic loads. More information is needed on the system's removal efficiencies and oxygen requirements during these periods. This is necessary to optimize designs for possible application to the direct treatment of combined sewer overflow wastewaters.

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*The complete report, entitled "Treatment of Municipal Wastewaters by the Fluidized Bed Bioreactor Process," (Order No. PB 88-140 280/AS; Cost: $25.95, subject to change) will be available only from:*

*National Technical Information Service*
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