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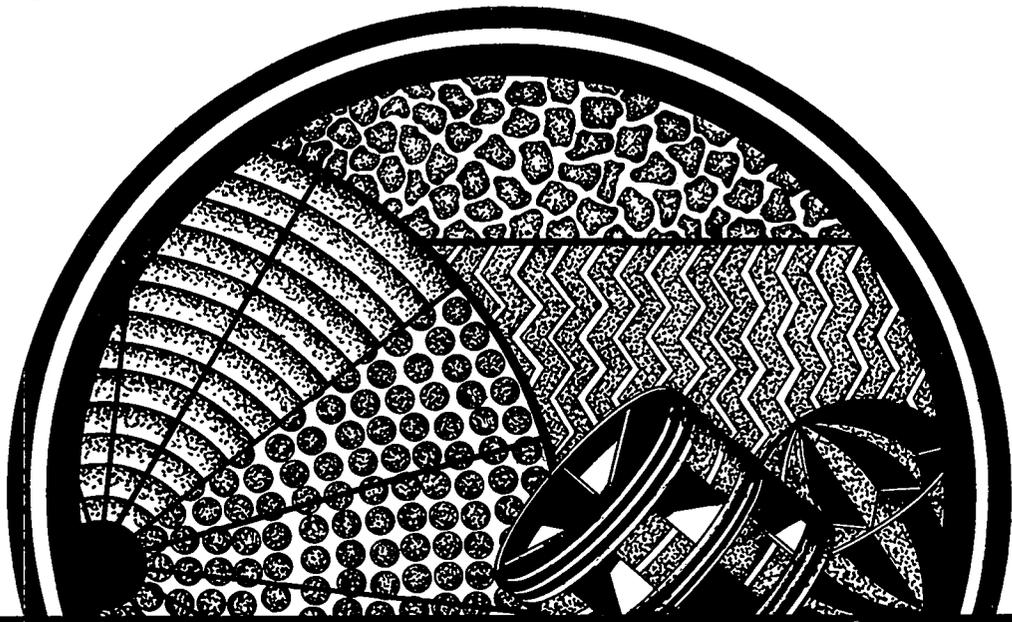


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Vol. III

ALCOHOL PRODUCTION WITH THE BACTERIUM ZYMOMONAS

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Clyde Engineering Service, care of J. Coveny

INTRODUCTION

At present there is plenty of petroleum, but conditions in the mid-East could change this overnight, so it would seem advisable to be prepared. *Zymomonas mobilis* ATCC 10988 has been found to attach readily to fibers such as cotton, Orlon and polyester. The bacteria are smaller than yeast, get in between the small fibers to present a large area to the sugar being pumped through, and react several times faster than yeast. Vigorous evolution of CO₂ in a vertical fermenter prevents plug flow, so horizontal fermenters are used. In a once through operation, pumping sugar slowly through stationary fibers, it took 25 hours to get an 80% yield, but when rotating the fibers, it took only 18 minutes residence time. Operation of an inclined, rotating fermenter is described, as well as several reactor configurations for commercial operation.

Laboratory runs were made using the apparatus shown in Fig. 1. It took about 24 hours to attach the bacteria to the fiber. There is a full page photo of *Zymomonas* on cotton in a recent periodical (1). The volume of the reactor is 550 ml. and the media was run through at 1800 ml/hr with a residence time of 18 minutes and an alcohol production of 133 gm/liter/hour, considerably higher than a survey made of other investigators in a recent periodical (2). Half of the weight goes to CO₂ and half to alcohol, so a 4% alcohol out the end represents an 80% yield. The 4% alcohol was compared with a standard 4% solution put in a gas chromatograph. Hallofil, a polyester from DuPont holds tight against the glass. A ridge on the top of the glass permits escape of CO₂. If rotation is too fast, the bacteria may break loose. ²LH Fermentation, Stoke Poges, Bucks, SL24EG, England, makes a device to measure cell adhesion by pumping media up through an orifice and then radially across the support. Dr. H. Fowler from the University College of Swansea, UK, made a report on this device at a recent meeting in Montreal.

Figures 2,3,4, and 5 show suggested commercial designs. In Fig. 2, the motor pulls the fiber one way and the spring pulls it back. When the fibers become clogged, chain 1 goes clockwise and chain 2 counter cw, so the striker bars interlock, flexing the fiber and dislodging biomass to the belt below. One problem in fermentations is that the cooling coils become coated with bacteria. Fig. 3 shows cooling water into the striker bars which are kept clean by rubbing against the fiber.

Fig. 4 shows a rotary fermenter. When the fiber becomes clogged, nozzles dislodge biomass to the screw conveyor below.

Fig. 5 shows a reactor filled with a polyester such as Hallofil. The sugar is pumped around and around until the alcohol starts to inhibit the reaction. Then it is pumped to the solvent extractor where a solvent such as dodecanol removes alcohol. The sugar settles to the bottom and is recycled to the reactor. Solvent extraction is much more energy efficient than distillation because of the water-alcohol azeotrope and the high heat of vaporization of water. When the fiber becomes clogged, the squeezer is

depressed to flex the fiber and dislodge excess biomass. This unit could be built small enough to mount on a truck which would operate in the north in the summer and go south in the winter. Patents are pending and available for license.

For the case where cellulose is converted to sugar and sugar to alcohol in one step with a bacterium such as *Clostridium thermocellum*, a flexible ceramic fiber is available.

A recent periodical (Ref.3) compares the different strains of *Zymomonas*; CP4 is better than 10988 (used in Clyde's experiments).

Methanol and ethanol can be decomposed to H₂ and CO to give more energy in a car or turbine which produce no pollution. The carburetor of a car produces a mist and the liquid tends to go straight, so it is difficult to get even distribution to the cylinders, using gasoline. A new design catalytic heat exchanger for decomposing methanol has been developed.

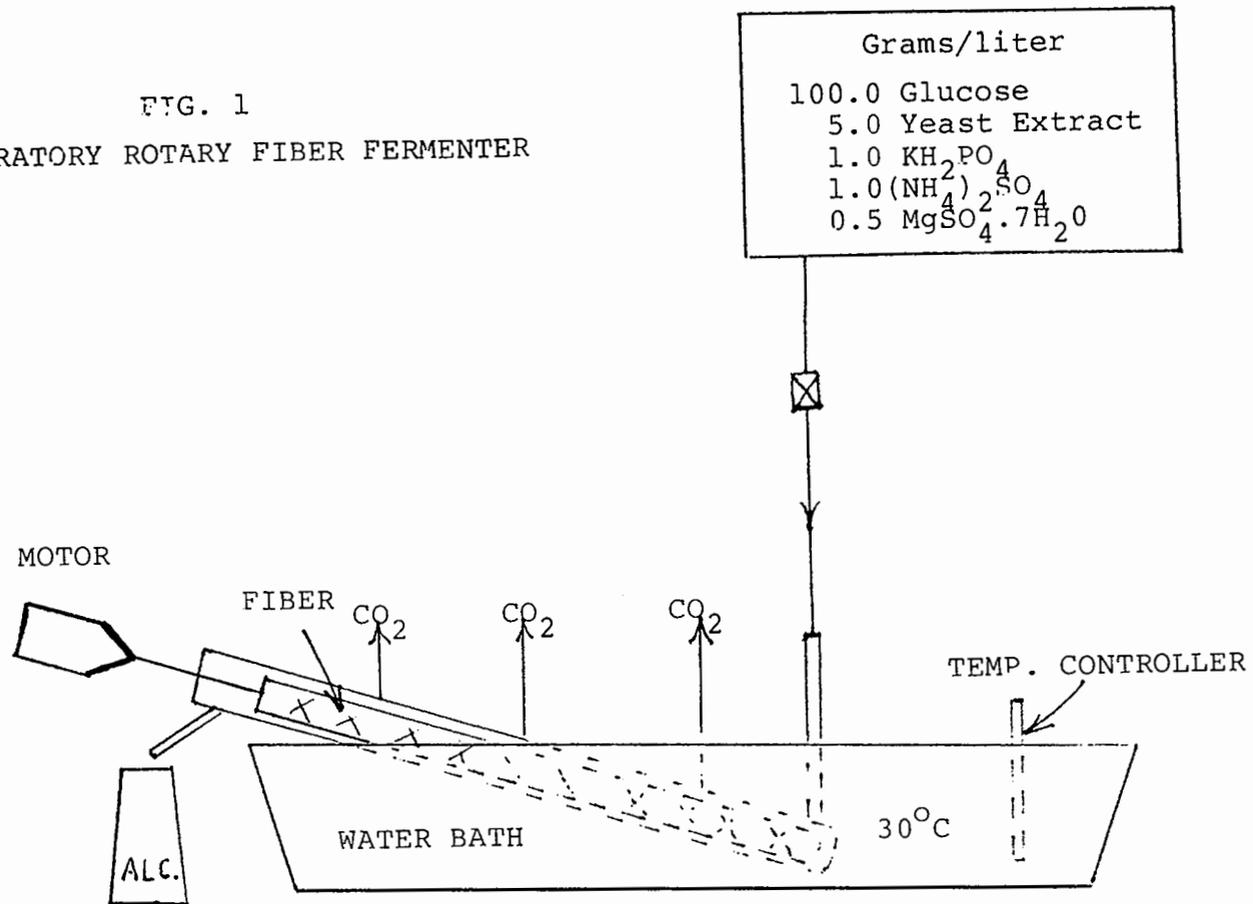
ACKNOWLEDGEMENTS

Thanks are due to Dr. Harvey Paige and student Mike Colvin of Alfred University Chemistry Dept. for gas chromatograph work and to Jim Wilkins of Rochester and Intertec Associates of Rochester, N.Y. for helpful advice.

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FIG. 1
LABORATORY ROTARY FIBER FERMENTER



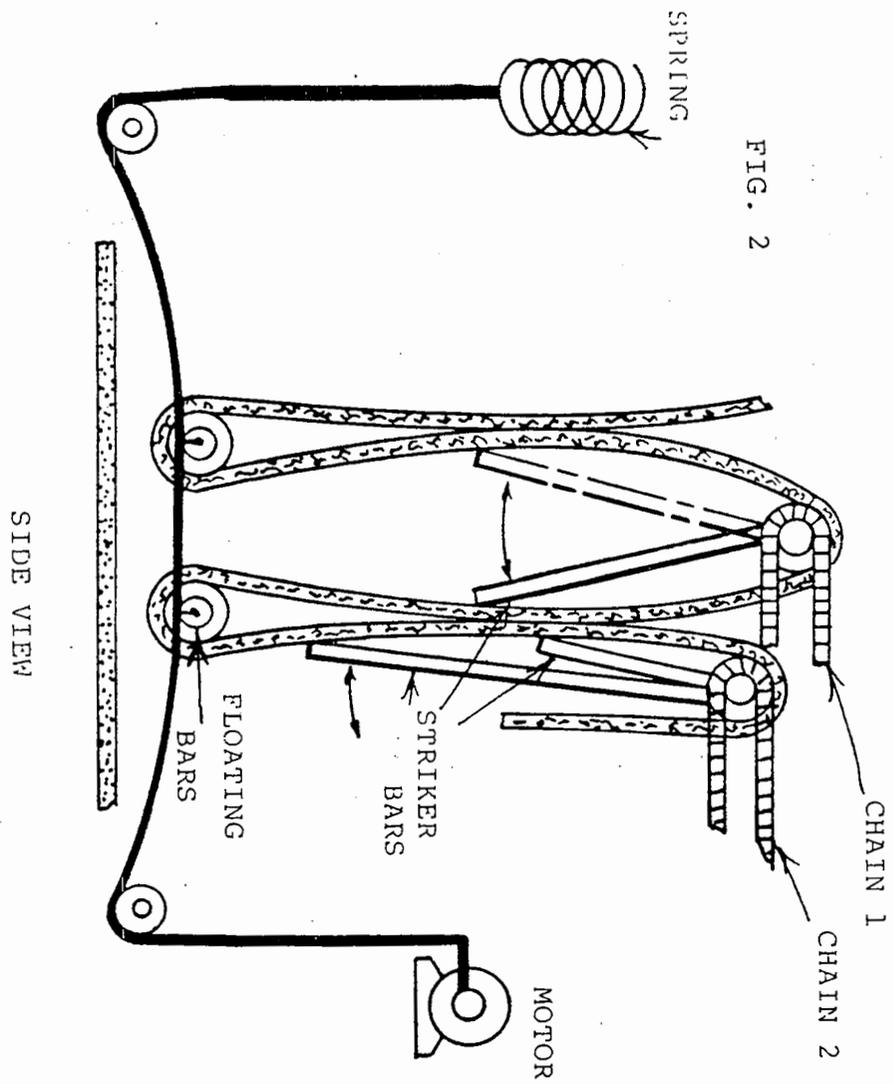
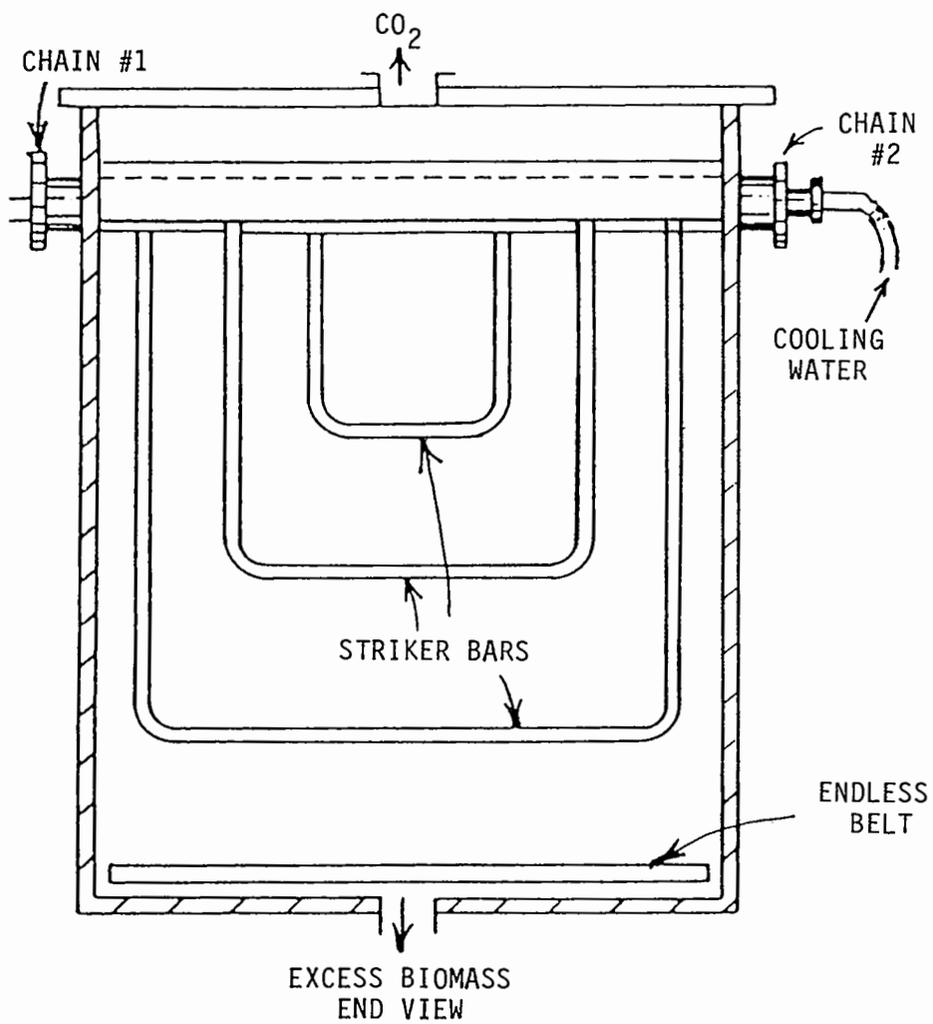


FIG. 3



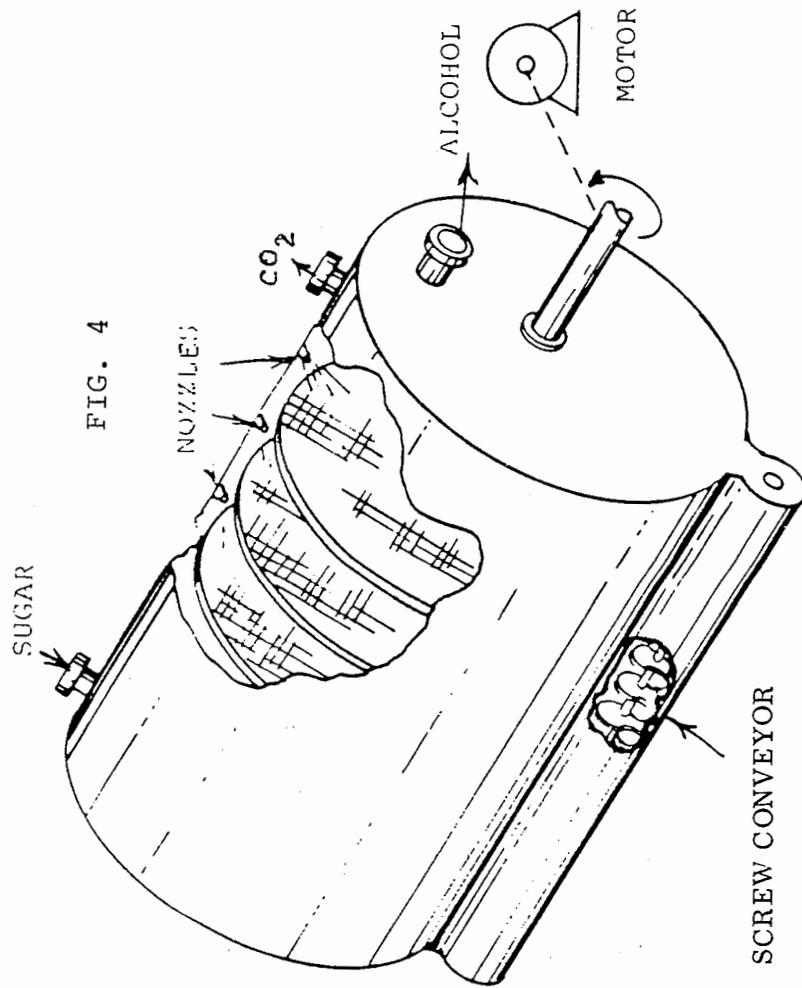
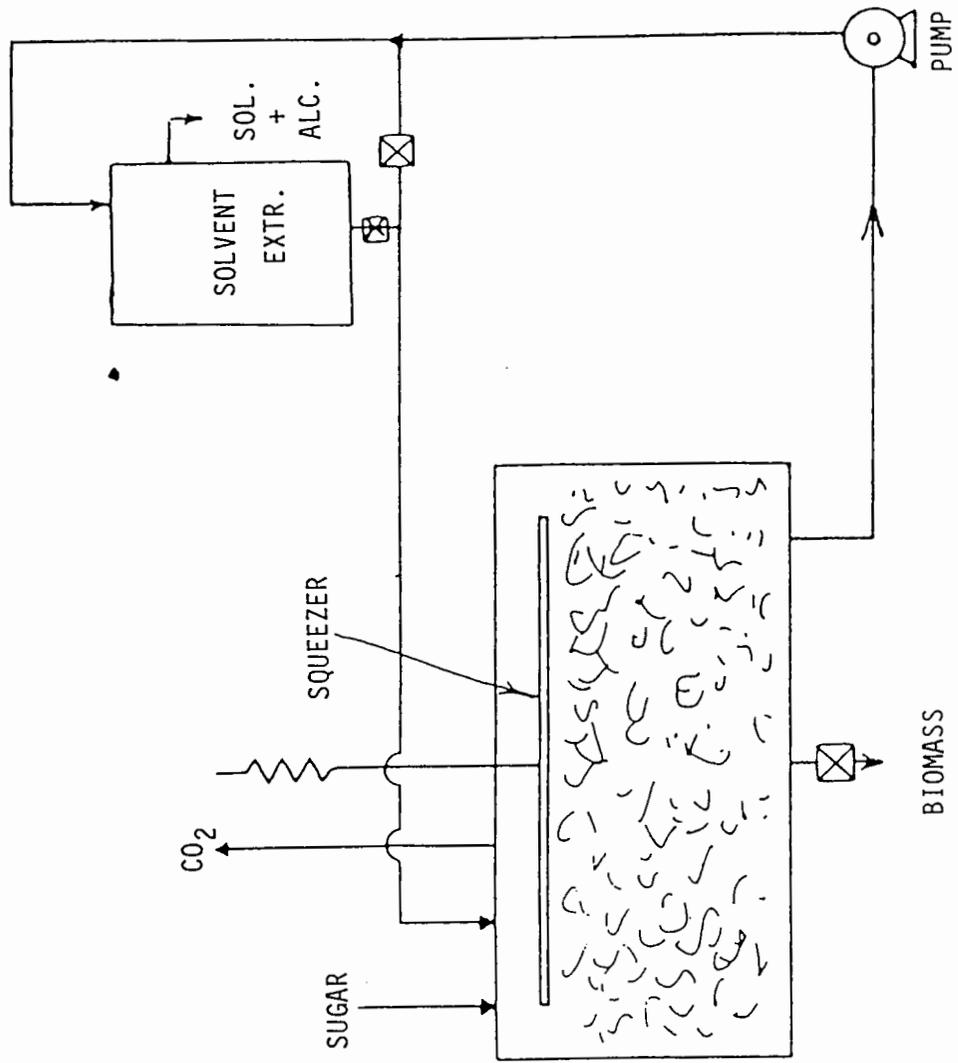


FIG. 5



DYNAMICS AND SIMULATION OF A BIOLOGICAL FLUIDIZED BED REACTOR

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INTRODUCTION

Getting consistently high efficiency for biological nitrification during cold weather is an important problem that was studied by operating and modeling a fluidized bed pilot plant. The pilot plant experiments, including some step change and impulse dynamic tests were encouraging, and a model has been used successfully to fit both steady-state and dynamic performance data.

THE FLUIDIZED BED PROCESS

The biological fluidized bed, developed for wastewater treatment in 1972 by Beer et al. (3), has been demonstrated by pilot plant and full-scale studies as being effective for aerobic and anaerobic treatment of domestic and industrial wastes to remove organic carbon and nitrogen (3, 14, 15, 22, 29).

The microorganisms necessary for treatment are grown as a film on small, dense carrier particles (usually sand) that are fluidized by an upward flow of wastewater. Because the

particles are small, the surface area of biofilm is very large. This creates the major advantage of the process -- the volumetric concentration of biomass may be ten times greater than an activated sludge process. Jeris et al. (14) reported bed volatile solids (BVS) concentrations as high as 14,200 mg/L for aerobic removal of BOD and 8500 mg/L for nitrification. The high BVS concentrations give high volumetric substrate removal rates and a correspondingly short hydraulic detention time. It also gives a high rate of oxygen consumption in an aerobic system. Empty-bed hydraulic detention times of 6 minutes for nitrification have been reported (14, 27), compared with at least two hours for a dispersed growth system (30).

The BVS are captive in the fluidized bed, recycle of biological solids is not needed to maintain a high solids concentration and a long mean cell retention time. These variables are maintained at effective levels by controlling the height of the fluidized bed. With time, the biofilm thickness increases and the particles become larger, less dense, and more buoyant, which causes the bed to expand more. Excessive expansion and excessive biofilm thickness are controlled by periodically removing the top portion of the fluidized bed, where the particles with the thickest biofilm tend to accumulate. These particles are removed and cleaned by shearing away the biofilm and returning the cleaned sand to the reactor. Except during this cleaning process the amount of suspended solids leaving the bed is negligible and it may be that a final clarifier is not needed. The biofilm sheared from the particles is collected as a dense sludge.

The small but growing literature on operating experience with fluidized bed treatment of many types of wastes has been summarized in an excellent review by Cooper and Wheeldon (6).

The first mechanistic model of a biological fluidized bed was by Mulcahy and LaMotta (20). Other modeling work has been published by Stathis (25) and Mulcahy et al. (21). Nutt et al. (22) studied the dynamic behavior of a fluidized bed for removal of organic carbon and modeled the results with time series methods.

THE PILOT PLANT

A schematic diagram of the pilot plant is shown in Figure 1. The 12-foot tall fluidized bed reactor was made of a 4-inch inside diameter plexiglass cylinder. Figure 2 is a photograph of the reactor. The influent was distributed into

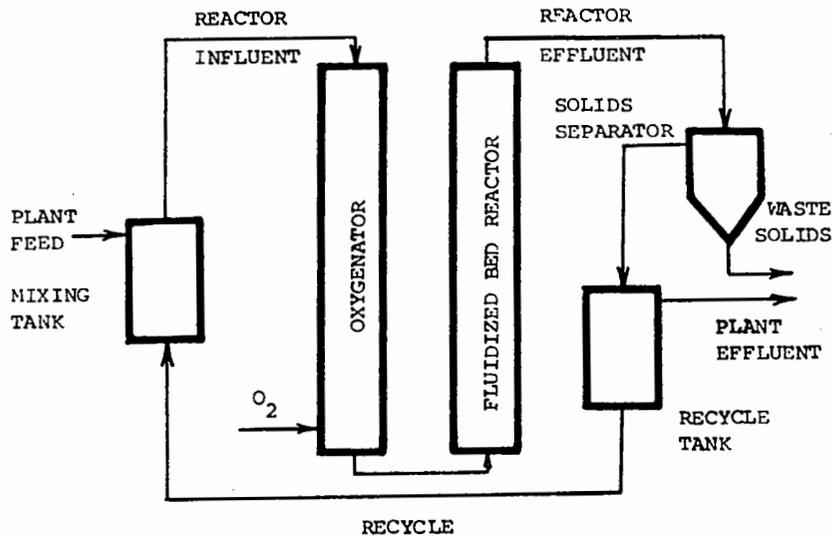


Figure 1 Pilot Plant Schematic



Figure 2 View of Oxygenator and Fluidized Bed Reactor

the reactor through a plate perforated by 25 evenly spaced, 1/32-inch diameter holes. Sampling taps were located 0.33, 1, 3, 5, 7, 9, and 11 ft above the inlet distribution plate. The reactor contained 33 lbs of river sand with an effective size (d_{10}) of 0.49 mm and a uniformity coefficient (d_{60}/d_{10}) of 1.32. The unfluidized depth of the sand bed was 5.4 feet.

The feed to the reactor was unchlorinated clarified activated sludge effluent that was diluted by recycle of fluidized bed effluent. The mixture was pumped at 30 psig through a 13-ft tall proprietary oxygenator provided by Dorr-Oliver Co. High purity oxygen was supplied to the oxygenator from 200 lb pressure cylinders. The feed to the reactor typically had a dissolved oxygen concentration of 50 to 55 mg/L and was delivered at a rate of 0.5-1.0 gpm.

PILOT PLANT OPERATION

Pilot plant performance stabilized at the beginning of November, 1980. A summary of operating data and waste characteristics is given in Table I. The short detention times and high biomass concentrations are typical of fluidized bed systems.

Grab samples were taken daily from August 15 through December 21 to characterize the pilot plant feed (the activated sludge plant effluent), the pilot plant effluent, and the mixed flow that was the influent to the reactor. These samples were analyzed for ammonium (NH_4^+ as N) and pH. All flowrates, the wastewater temperature, and the reactor effluent dissolved oxygen were recorded daily. Data for November 1 to December 21, 1980 are given in Table II.

PILOT PLANT PERFORMANCE: STEADY STATE

Complete (>99 percent) ammonia removal was first observed on November 1, 1980, and from then until the plant was shut down on December 22, removal efficiencies were consistently above 90 percent, except when too little oxygen had been supplied. This is shown by Table II.

The consistency of the process is more apparent if the amount of ammonia removed, rather than percent removed, is considered. The amount of ammonia that could be consistently removed was about 11.0 mg/L over a range of conditions. Figure 3 shows the reduction in ammonia concentration plotted against the feed concentration. The feed concentration of ammonia that can be satisfied by the oxygenator under normal conditions (supply of about 50-55 mg/L at 30 psig and flow

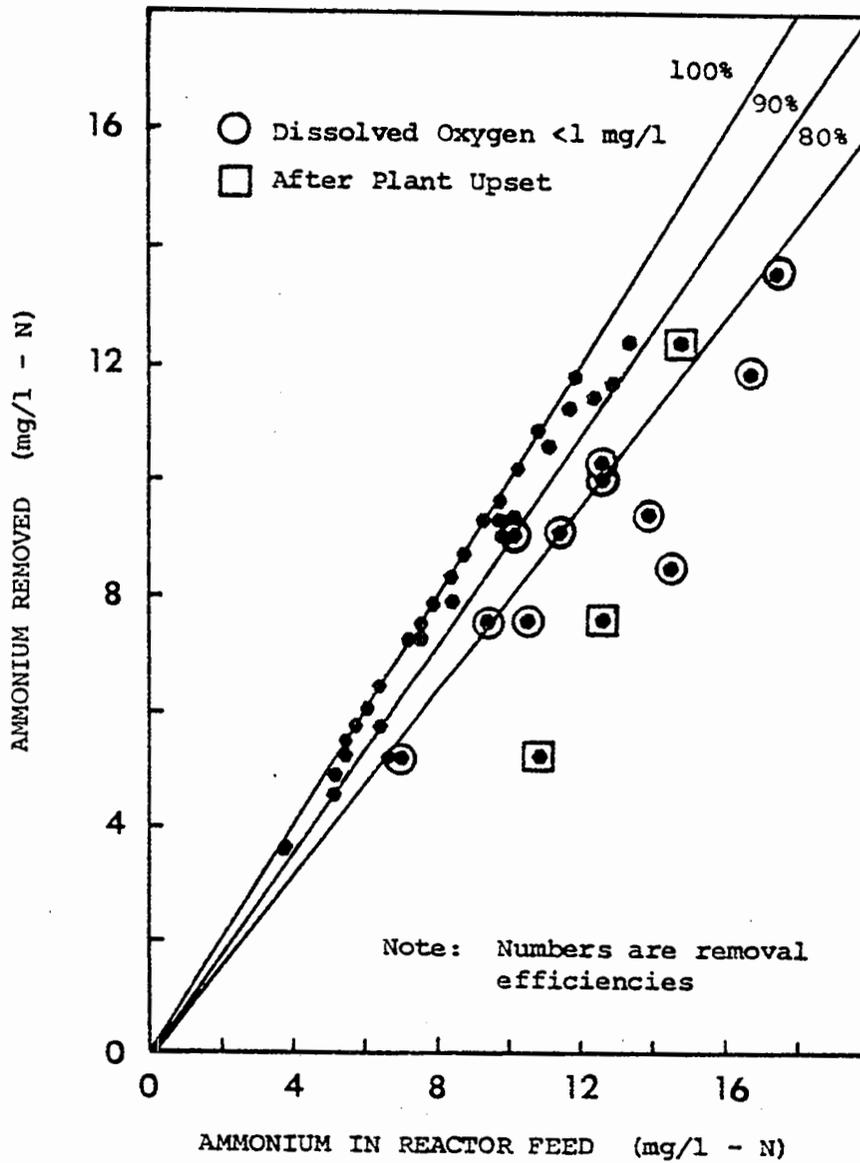


Figure 3 Reactor Ammonium Removal vs.
 Feed Concentration -- Nov. 1 to
 Dec. 21, 1980

Table I. Summary of Operating Conditions and Influent Characteristics

<u>Parameter</u>	<u>Range</u>
Hydraulic Flux	
Nov. 1 - Dec. 13	4.6 - 7.0 gpm/ft ²
Dec.13 - Dec. 22	9.0 - 9.2 gpm/ft ²
Hydraulic Retention Times (based on empty bed volume)	6.5 - 14.0 Min.
Bed Volatile Solids	10 - 12,000 mg/l
Recycle Ratio (R/Q)	0-1
Temperature	12 - 20°C
pH	
Influent	7.2 - 7.7
Effluent	6.8 - 7.2
Oxygen Flow	50 - 150 mg/min
(NH ₄ ⁺ -N)	6 - 21 mg/l

of 0.5-1.0 gpm) is 11-12 mg/l, which is marked on the abscissa. When the influent ammonia concentration was equal to or less than this the removal efficiency was 90 to 100 percent, often 100 percent. When the feed concentration exceeded 11-12 mg/L, the process continued to remove about that amount (it could remove more with a larger supply of oxygen). In Figure 3, circles are drawn around the points for which the effluent dissolved oxygen concentration was less than 1 mg/L at the time of sampling. Most of the occasions showing poor removal can be explained by oxygen limitation.

There were two reasons for having too low a dissolved oxygen concentration in the effluent. One was a pressure imbalance in the gas and liquid flow to the oxygenator that resulted when the reactor inlet distribution plate would become fouled. The other reason was simply the nitrification oxygen demand exceeding the oxygen transfer capacity of the equipment.

On a few days, the liquid flow stopped or operation was otherwise disrupted by equipment problems. Grab samples taken from 15 to 60 minutes after re-starting the plant showed that the fluidized bed quickly regained its normal

Table II Daily Fluidized Bed Operating Conditions

Date (1980)	Flow (gpm)	Bed Depth (ft)	Det. Time (min)	D.O.** (mg/l)	Ammonium Nitrogen Inf.† (mg/l)	Ammonium Nitrogen Feed‡ (mg/l)	Removal Eff.††† (mg/l)	Removal Efficiency
11/1	0.64	6.07	6.19	7.00	14.31	6.43	0.03	1.00
2	0.48	6.10	8.30	0.00	14.36	5.33	0.04	0.99
3	0.54	6.10	7.38	0.30	17.35	9.40	2.03	0.78
4	0.55	6.10	7.24	5.00	14.03	7.18	0.00	1.00
5	0.50	6.04	7.38	16.00	14.03	5.01	0.07	0.99
6	0.52	6.14	7.70	16.00	13.20	3.68	0.00	1.00
7	0.64	6.23	6.36	0.70	13.99	6.50	1.35	0.79
8	0.45	6.17	8.95	0.40	13.18	8.45	0.00	1.00
9	0.51	6.30	8.06	12.50	16.10	7.44	0.00	1.00
10	0.50	6.23	3.14	0.00	16.42	14.33	5.82	0.59
11	0.47	6.30	3.75	17.00	12.38	5.28	0.04	0.99
12	0.50	6.50	9.48	2.50	13.11	13.11	0.59	0.95
13	0.51	6.14	7.85	0.00	13.96	6.01	1.73	0.74
14	0.48	6.30	8.57	0.00	15.72	11.41	2.38	0.79
15	0.53	6.14	7.56	11.00	18.27	3.34	0.48	0.94
16	0.51	6.30	8.06	0.50	19.79	10.59	3.08	0.71
17	0.52	6.40	8.03	25.00	20.65	6.34	0.56	0.91
18	0.48	6.66	9.06	25.00	14.76	5.07	0.63	0.88
19	0.49	6.76	9.00	0.00	15.51	10.33	1.39	0.87
20	0.45	6.76	9.30	4.00	14.17	9.88	0.58	0.94
21	0.46	6.96	9.87	0.00	17.49	17.49	4.30	0.77
22	0.64	7.05	7.19	1.00	19.20	12.42	1.11	0.91
23	0.55	7.12	8.45	1.00	20.85	13.92	4.63	0.67
24	0.43	7.02	10.66	0.00	19.97	11.83	0.74	0.94
25	0.47	7.09	9.84	0.00	15.95	12.73	2.32	0.78
26	0.58	7.32	8.23	4.20	16.45	11.07	0.65	0.94
27	0.63	7.58	7.31	25.00	16.03	7.67	0.35	0.95
28	0.61	7.22	7.72	1.40	15.30	7.49	0.14	0.98
29	0.60	7.22	7.35	0.60	15.54	9.84	0.74	0.92
30	0.62	7.25	7.53	25.00	13.79	6.11	0.11	0.98
12/1	0.50	6.76	3.32	0.00	11.73	11.31	0.07	0.99
2	0.52	6.86	8.61	25.00	13.16	5.89	0.00	1.00
3	0.53	6.86	8.45	25.00	11.55	6.20	0.00	1.00
4	0.48	6.89	9.37	0.00	12.57	12.50	2.38	0.81
5	0.45	6.79	9.85	25.00	13.76	7.93	0.00	1.00
6	0.46	6.82	9.68	7.50	16.28	10.12	0.00	1.00
7	0.52	7.05	8.85	0.00	17.71	16.88	5.32	0.68
8	0.35	7.22	13.46	0.00	19.32	12.86	1.07	0.92
9	0.39	7.32	12.45	18.00	15.72	10.68	0.00	1.00
10	0.44	7.81	11.38	25.00	15.58	9.29	0.00	1.00
11	0.46	7.94	11.27	25.00	13.55	7.92	0.00	1.00
12	0.44	7.78	11.54	25.00	14.43	5.60	0.00	1.00
13	0.85	9.06	6.95	19.00	11.62	11.62	0.00	1.00
14	0.88	9.09	6.74	5.50	12.71	14.60	2.24	0.85
15	0.79	8.89	7.35	0.00	15.32	12.67	5.36	0.60
16	0.83	9.09	7.15	25.00	12.99	10.77	5.53	0.49
17	0.81	9.22	7.43	2.00	11.62	11.13	0.63	0.94
18	0.78	9.65	9.07	19.50	8.99	9.70	0.00	1.00
19	0.78	10.10	3.56	14.50	8.82	9.70	0.32	0.97
20	0.78	10.20	8.54	0.00	8.54	8.79	0.00	1.00
21	0.80	10.50	8.57	18.00	5.95	5.74	0.00	1.00

* Based on empty bed volume

** Dissolved Oxygen

† Pilot plant influent

‡ Reactor feed

††† Reactor effluent

efficiency. The exception was an upset on December 14 for which the slow recovery cannot be explained. The data that correspond to these disruptions are highlighted with boxes in Figure 3.

Temperature decreased from 20°C to 12°C as the experiment progressed. This seemed to have little or no effect on performance. Temperature effects in fluidized bed systems have been reported to be less critical than in dispersed growth systems (6).

The depth of the fluidized bed over the course of the experiment, at the indicated hydraulic fluxes, is plotted in Figure 4. At the end of the study the expanded bed depth was increasing steadily and the thickness of the biofilm was increasing. Early in the study there was no visible biofilm, yet complete ammonia removal was achieved. During the experiment no sand was removed or cleaned.

PILOT PLANT PERFORMANCE: DYNAMIC EXPERIMENTS

Three types of dynamic experiments were performed: (1) step forcing of ammonia concentration, (2) step forcing of hydraulic flux, and (3) impulse forcing of ammonia and nitrite concentrations.

Step Forcing of Ammonia

A step forcing of ammonia from 13 mg/L to 22 mg/L was done by adding NH_4Cl solution at a constant rate to the feed reservoir. The total flow rate was unchanged and the empty-bed hydraulic detention was 7.8 minutes. The response observed is shown in Figure 5. Essentially all of the additional 9 mg/L ammonia broke through the bed; the process could not remove more than the 12-13 mg/L that was being removed before the step change was made. Effluent dissolved oxygen dropped from 2.5 to less than 0.5 mg/L when ammonia in the effluent started to increase. The amount of ammonia removed is the stoichiometric equivalent of the available oxygen.

Step Forcing of Flow

The hydraulic flux was increased suddenly from 4.6 gpm/sq.ft to 9.2 gpm/sq.ft; this reduced the empty-bed detention time from 12.4 to 7.3 minutes. Figure 6 shows the response. There was little ammonia breakthrough and only a small amount

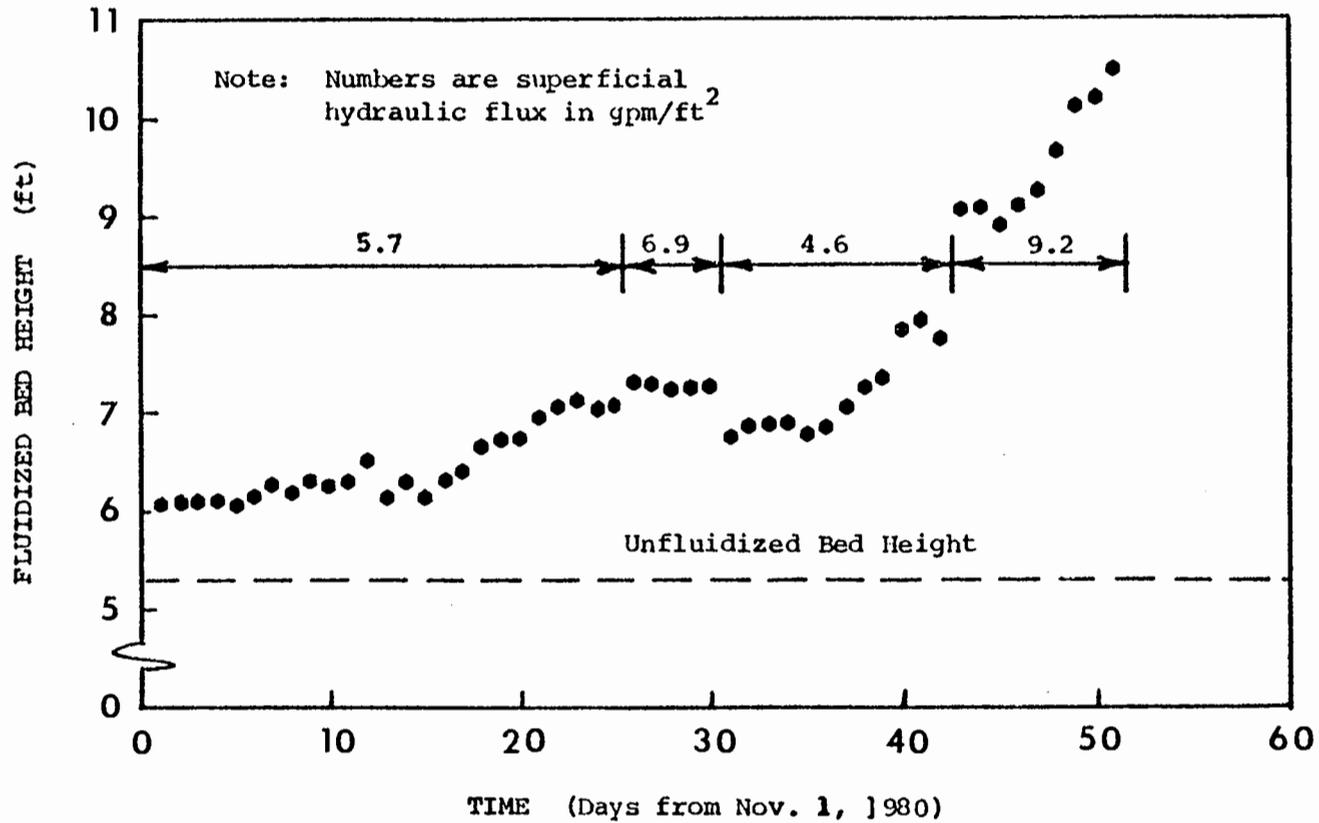


Figure 4 Height of Fluidized Bed During Pilot Plant Study
Nov. 1 to Dec. 21, 1980

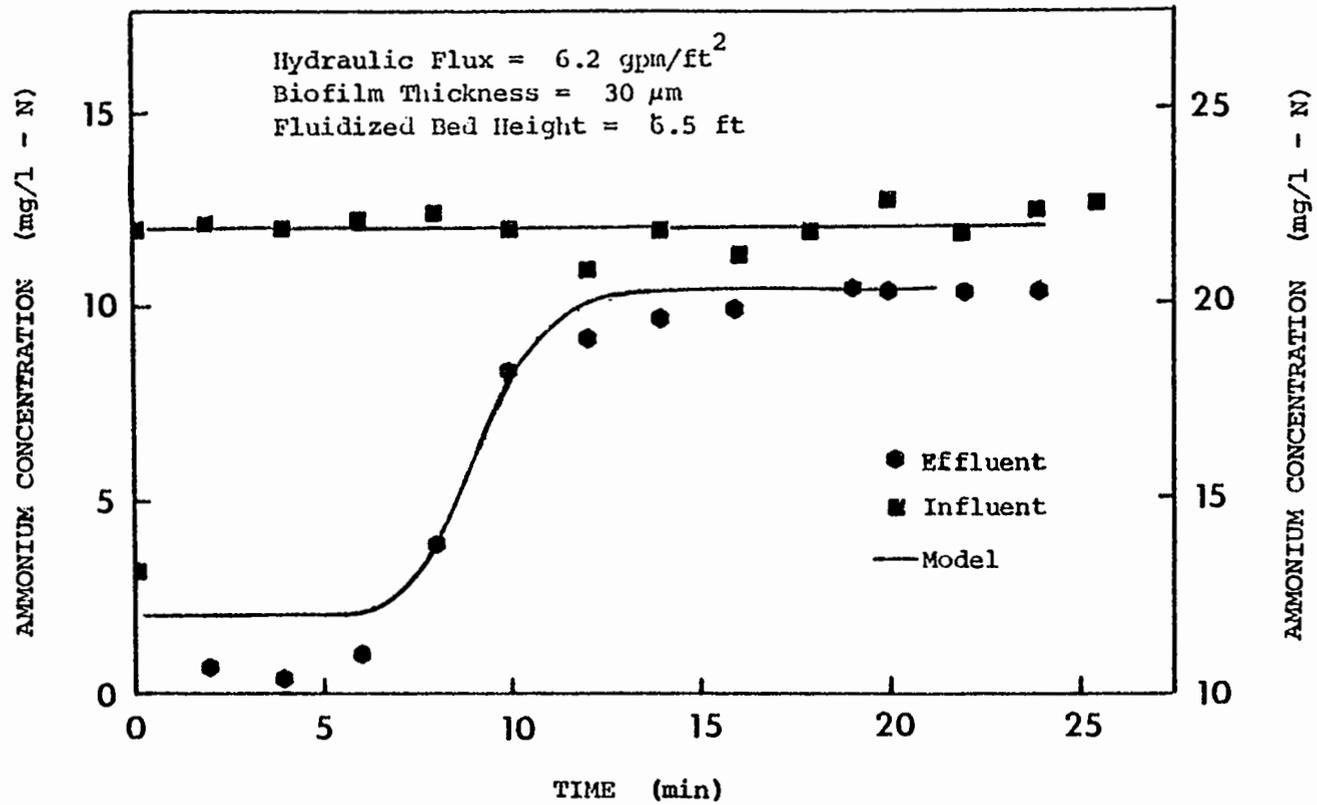


Figure 5 Experimental and Simulated Response to Step Forcing of Influent Ammonium Concentration

of nitrite was observed in the effluent. The influent ammonia concentration was 11.5 to 12.0 mg/l during the experiment. The bed absorbed a doubling in flow rate without deterioration of effluent quality. How this can occur is seen from the profiles shown in Figure 7. Before the step change, the influent ammonia was being removed in the first few feet of the bed. The upper portion of the bed was available to take up fluctuations in the influent. Figure 7 shows, however, that the profile after the step change was not much different and the ammonia was still removed in half the bed.

Impulse Forcing of Ammonia and Nitrite

At a hydraulic flux of 9.2 gpm/sq.ft (empty bed detention time of 7.84 minutes), a pulse containing 185 mg of ammonia (NH_4^+) and 185 mg of nitrite (NO_2^-), both measured as nitrogen, was instantaneously injected into the inlet of the pilot reactor. Before injection of the pulse the influent ammonia concentration was 9.4 mg/L and the nitrite concentration was less than 0.5 mg/L; the quality of the feed during the experiment was steady. The response of the effluent to the impulse is shown in Figures 8 and 9. The response curves were integrated numerically and it was found that 52.6 percent of the injected ammonia and 11.1 percent of the injected nitrite were removed.

Later in the paper this and additional data will be fitted with the mathematical model presented in the following section.

SUMMARY OF OBSERVED RESULTS

At the low temperature of 12°C, six to ten minutes hydraulic detention time removed 11-12 mg/L ammonia, provided the process had sufficient dissolved oxygen. Recycle at a 1:1 ratio was needed to dilute the feed to this range since the available oxygen supply was 50-55 mg/L. The process was stable and recovered quickly from shutdowns. Under normal operating conditions all ammonia removal took place in less than half the bed volume. The remaining upper portion of the bed was able to absorb sudden changes in feed concentration and flow rate. These results held true even when the biofilm was invisible.

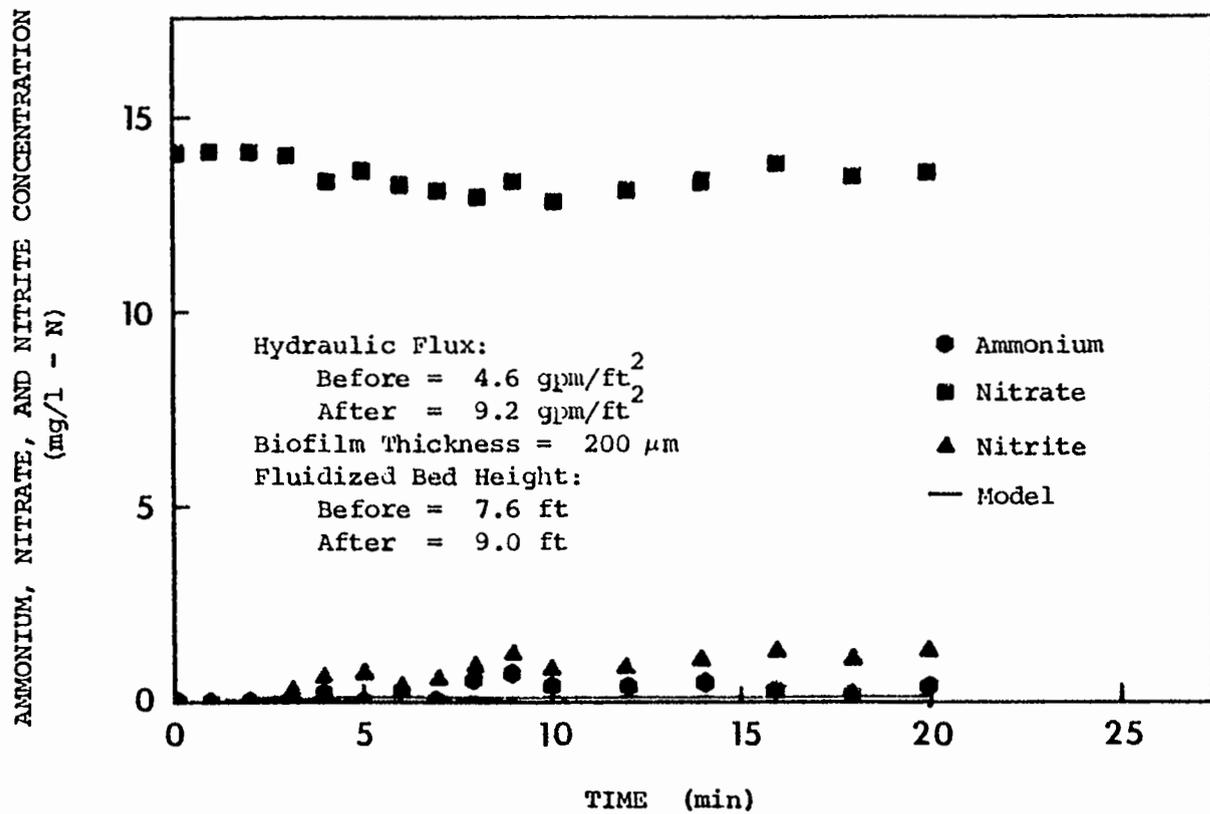


Figure 6 Experimental and Simulated Responses to Step Forcing of Hydraulic Flux

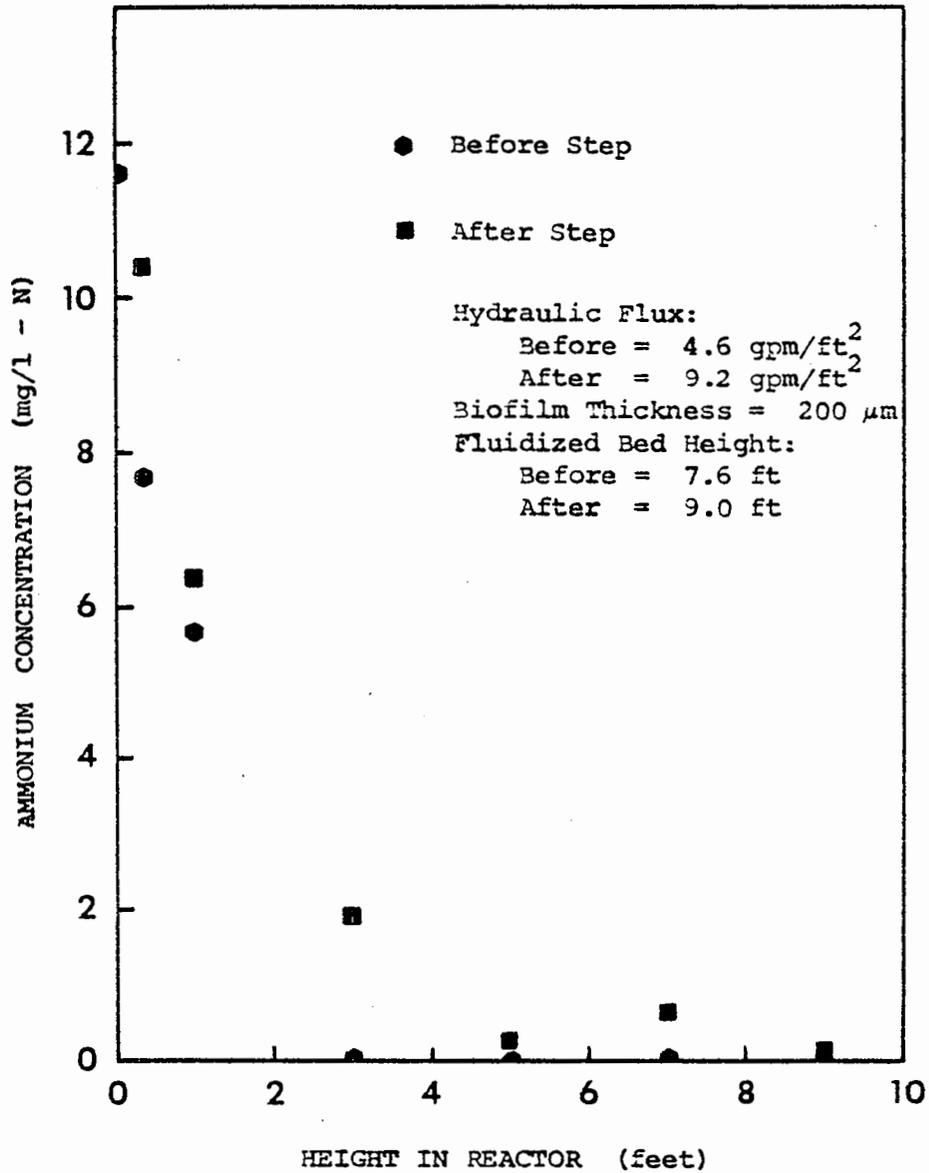


Figure 7 Reactor Ammonium Profiles
 Before and After Step Forcing
 of Hydraulic Flux

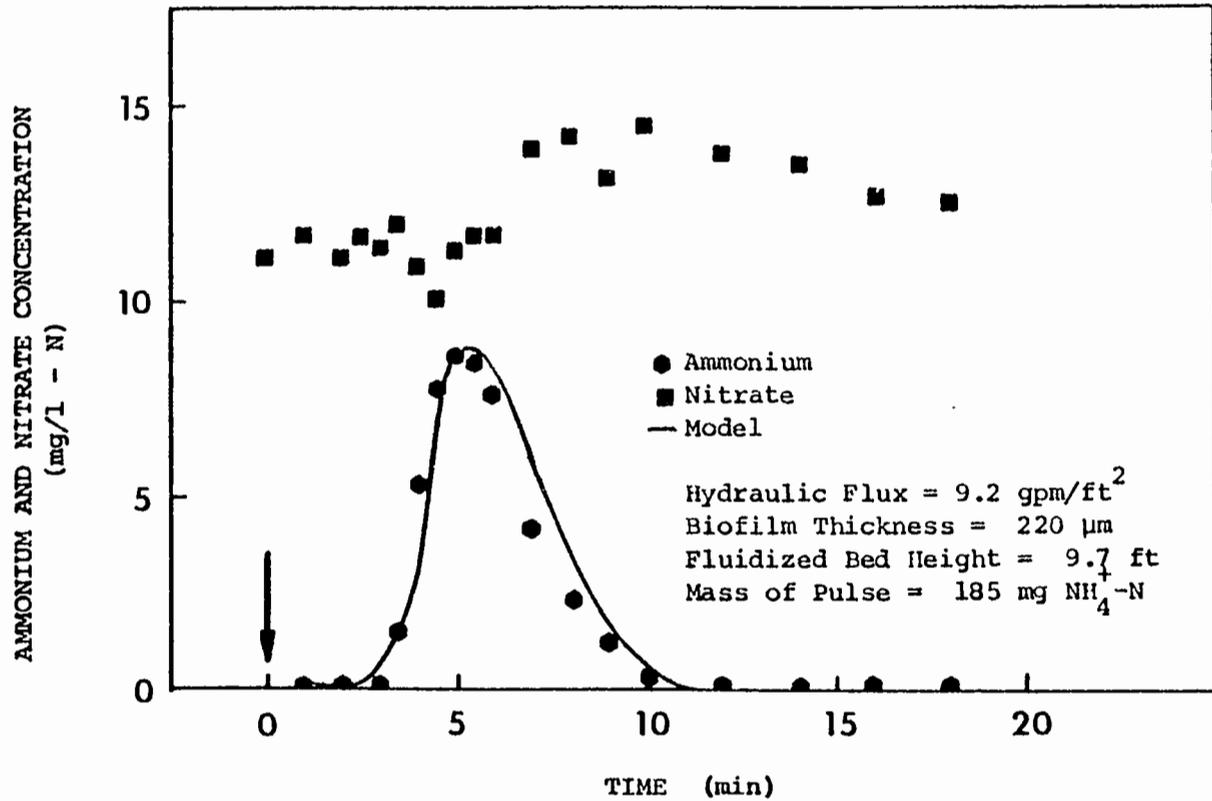


Figure 8 Experimental and Simulated Responses to Impulse Forcings of Ammonium and Nitrite Concentration

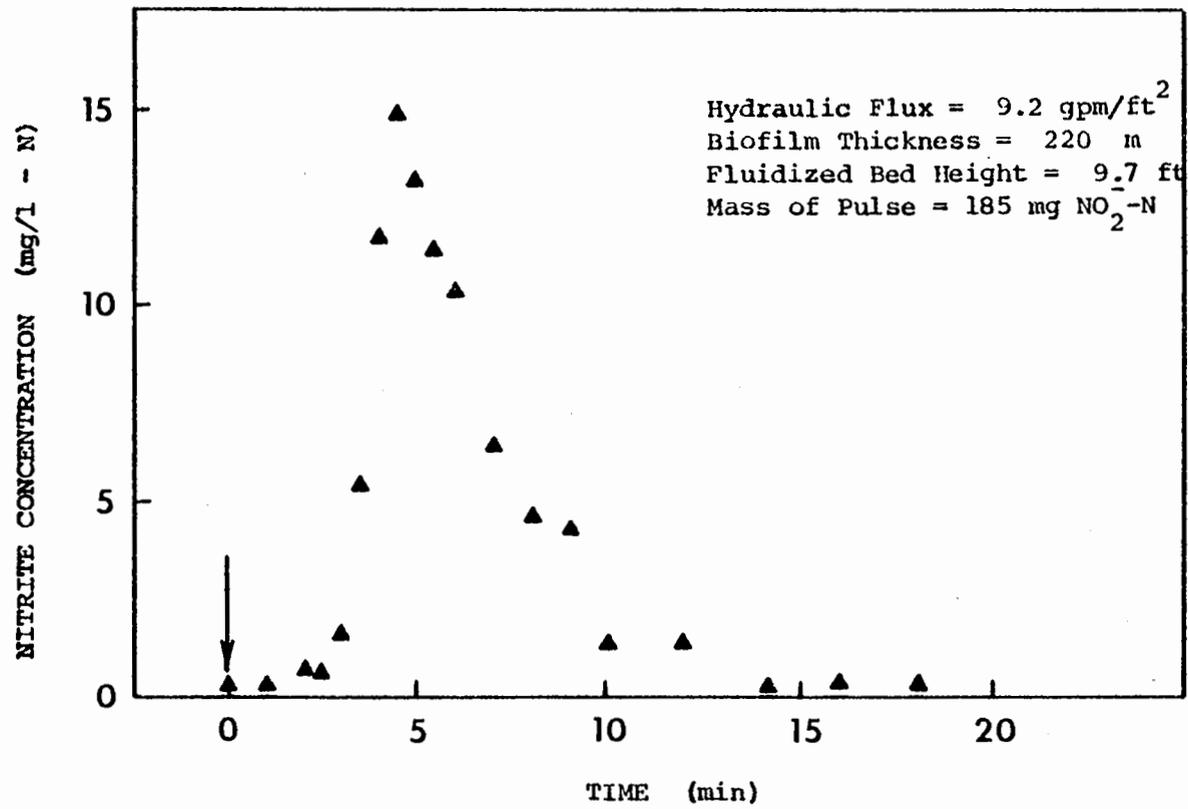
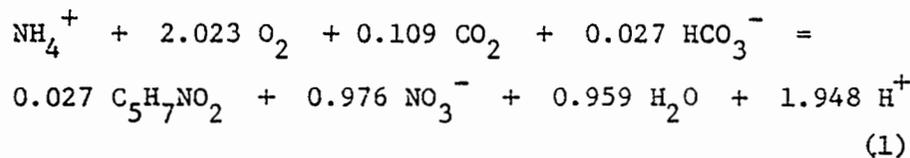


Figure 9 Experimental Response to Impulse Forcings of Ammonium and Nitrite Concentration

BIOFILM MODEL

Reaction Stoichiometry

Nitrification is carried out in two steps by two genera of microorganisms. Nitrosomonas oxidizes ammonium (NH_4^+) to nitrite (NO_2^-) and Nitrobacter oxidizes the product nitrite to nitrate (NO_3^-). Using the method of Christensen and McCarty (5), for a sludge age of 6 days, decay coefficient 0.12 days^{-1} , and an ATP transfer efficiency of 0.6, the overall stoichiometric equation for the reaction is:



Biofilm Material Balance

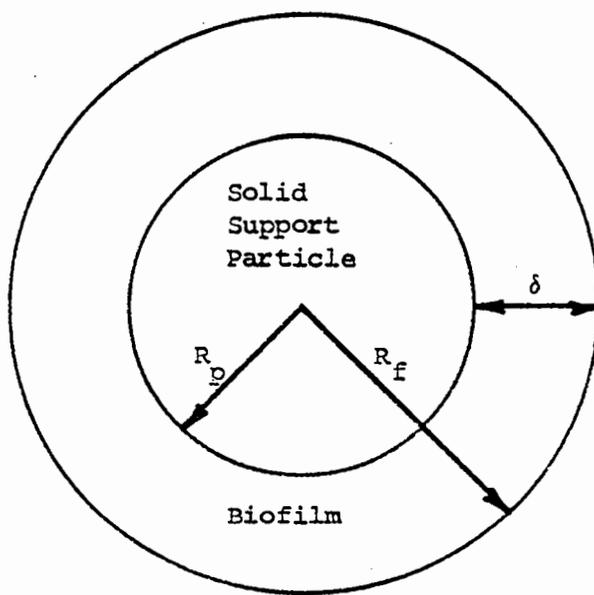
The model of the biological catalyst particle describes the physical transport of reacting species (NH_4^+ , HCO_3^- , O_2 , and H_2CO_3) into and through the biofilm and the kinetics of consumption and production of the species.

Liquid to solid-phase mass transfer in biofilm systems has been likened to heterogeneous chemical catalysis (1). The catalyst particle is assumed spherical. Its center, which may be most of its volume, is inert. Reactions occur only in the thin film surrounding the inert carrier particle. Flux into the biofilm is considered to be controlled by (1) mass transport of the chemical species from the liquid into the biofilm and (2) substrate removal resulting from microbial growth and decay. Mass transport is described by Fick's first law of diffusion using an effective diffusivity that combines molecular diffusion and pore convection.

The bioparticle is shown schematically in Figure 10. The continuity equation for this geometry, assuming constant density and diffusivity is (4):

$$\frac{\partial C}{\partial t} = D_e \frac{1}{r^2} \frac{\partial}{\partial r} \left(r^2 \frac{\partial C}{\partial r} \right) + R(C) \quad (2)$$

with the boundary conditions: $C = C_b$ at $r = R_f$ and $\partial C / \partial r = 0$ at $r = R_p$, and initial conditions $C(r,t) = C_f(r,0)$ at $t = 0$. This assumes that there is no resistance to external mass transfer from the bulk solution to the particle and that



Bulk Solution

Figure 10 Schematic of Fluidized Bed Bioparticle

the carrier particle is non-porous so that there is no mass transfer into the carrier.

This continuity equation can be solved for the transient or steady state concentration profiles in the film and for the surface flux for a given particle geometry and reaction rate expression $R(C)$. Analytical solution is possible if the expression for $R(C)$ is simple, such as zero or first order kinetics. For nonlinear kinetic functions analytical solutions of the continuity equation are not possible and numerical solutions must be used.

The continuity equation is written for each species that can limit the reaction rate: ammonia, dissolved oxygen, bicarbonate, and carbonic acid. Thus, four simultaneous equations are used to describe the biofilm.

Microbial Growth and Ammonium Conversion

The rate of substrate conversion in the biofilm is related to the rate of bacterial growth. If more than one species can be growth limiting, the fundamental model of Monod (19) is modified. In the fluidized bed, the cell growth model, including cell decay, is:

$$\left(\frac{dX_n}{dt}\right)_f = X_n \left(\hat{\mu}_n \frac{(NH_4)}{K_s^n + (NH_4)} \frac{(O_2)}{K_s^o + (O_2)} - b\right) \quad (3)$$

Ammonium utilization is related to cell growth by the yield coefficient:

$$Y_n [=] \frac{\text{mass of volatile solids produced}}{\text{mass of substrate consumed}}$$

and:

$$\begin{aligned} \left(\frac{d(NH_4)}{dt}\right)_f &= - \frac{1}{Y_n} \left(\frac{dX_n}{dt}\right)_g \\ &= - \frac{X_n \hat{\mu}_n}{Y_n} \frac{(NH_4)}{K_s^n + (NH_4)} \frac{(O_2)}{K_s^o + (O_2)} \end{aligned} \quad (4)$$

This is the function $R(C)$ used in the continuity equation (Equation (2)) for the flux of ammonia into the biofilm.

Oxygen Utilization

Oxygen is required in the approximate amounts of 4.5 mg O₂/mg N for nitrification, 1 mg O₂/mg BOD oxidized, and 1.42 mg O₂/mg biomass oxidized for microbial respiration. The rate of oxygen consumption in the biofilm is modeled as the sum of the three removal mechanisms:

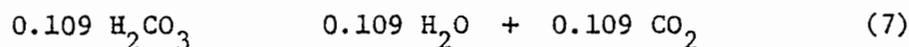
$$\begin{aligned} \left(\frac{d(O_2)}{dt}\right)_f = & - \frac{X_n \hat{\mu}_n}{Y_o} \frac{(NH_4)}{K_s^n + (NH_4)} \frac{(O_2)}{K_s^o + (O_2)} \\ & - \frac{X_b \hat{\mu}_b}{Y_b} \frac{(BOD)}{K_s^b + (BOD)} - 1.42 b (X_n + X_b) \end{aligned} \quad (5)$$

This kinetic term is used in Equation (2) to form the continuity equation for dissolved oxygen.

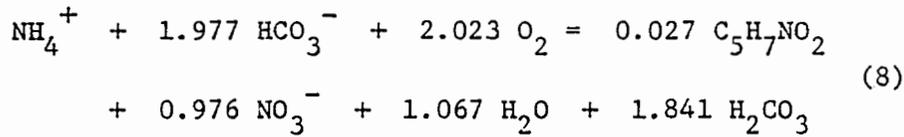
Carbonate System

The stoichiometry of nitrification shows that for each mole of NH₄⁺ consumed, 0.027 moles of HCO₃⁻ and 0.109 moles of CO₂ are consumed, and 1.95 moles of H⁺ are produced. If substantial amounts of alkalinity are destroyed, the pH might drop to a level that inhibits nitrification. Calculation of concentration profiles for alkalinity and H⁺ in the film are included in the model so the growth rate can be adjusted if inhibitory pH levels should occur. In most natural waters and municipal wastewater, the carbonate system is the principal pH buffer (31) - no other buffering is included in the model.

The interactions between the species in the carbonate system are described by assuming equilibrium always exists (i.e., the reactions occur much faster than nitrification) and are modeled as follows:



These are combined with the nitrification reaction (Equation (1)) to give:



It is assumed that HCO_3^- is removed and H_2CO_3 is produced only in the nitrification process. The kinetic equations for these species are:

$$\left(\frac{d(\text{HCO}_3^-)}{dt}\right)_f = - \frac{X_n \hat{\mu}_n}{Y_h} \frac{(\text{NH}_4)}{K_s^n + (\text{NH}_4)} \frac{(\text{O}_2)}{K_s^o + (\text{O}_2)} \quad (9)$$

$$\left(\frac{d(\text{H}_2\text{CO}_3)}{dt}\right)_f = \frac{X_n \hat{\mu}_n}{Y_{h_2}} \frac{(\text{NH}_4)}{K_s^n + (\text{NH}_4)} \frac{(\text{O}_2)}{K_s^o + (\text{O}_2)} \quad (10)$$

Each of these equations is used as $R(C)$ in the continuity equation (Equation (2)) and solved for the concentration profile and surface flux of each species.

Once these species are determined, the pH at each point in the biofilm can be estimated from the equilibrium reaction:



which has a stability constant K_a defined by:

$$K_a = \frac{(\text{HCO}_3^-) (\text{H}^+)}{(\text{H}_2\text{CO}_3)} \quad (12)$$

Typical values for K_a in natural waters at 20°C are $10^{-6.25}$ to $10^{-6.35}$. Solving equation (12) for pH gives:

$$\text{pH} = \text{p}K_a - \log_{10} \left(\frac{(\text{H}_2\text{CO}_3)}{(\text{HCO}_3^-)} \right) \quad (13)$$

This pH value is then used to modify the reaction rate as detailed in the next section. The pH is calculated at each point in the film so that local reaction rates can be estimated.

Dependence of Kinetic Parameters on pH and Temperature

Below pH 6.0, nitrification is severely limited. As pH increases, the rate increases to a maximum at pH 8.6 - 9.0, above which the rate decreases (30). This pH dependence is usually expressed by a modification of the maximum growth rate $\hat{\mu}_n$. Huang and Hopson (13) studied this relationship in a fixed growth system. The empirical approximating function:

$$\frac{\hat{\mu}_n}{\hat{\mu}_{n\max}} = 0.22 + 0.39 (\text{pH}-6.)^2 - 0.14 (\text{pH}-6.)^3 + 0.012 (\text{pH}-6.)^4 \quad (14)$$

adequately fits their data over the pH range of 6.0 to 9.0.

The maximum specific growth rate of nitrifying bacteria changes with temperature over the range of 5 - 35°C, common to the natural environment (7,8,9,13,30). Most investigators found an optimum temperature of about 30°C in both attached and suspended growth systems. Downing et al. (7, 9) represented the dependence by an exponential term of the form:

$$\hat{\mu}_t = 0.0196 \exp (0.098 (T - 15)) \quad (15)$$

with $\hat{\mu}_t$ in hours⁻¹ and T in degrees C.

The half-saturation constant for nitrogen, K_s^n is also a function of temperature (7, 9):

$$K_s^n = 10^{(0.051 T - 1.158)} \quad (16)$$

with K_s^n in mg/l - N and T in degrees C. K_s^0 has been assumed temperature independent.

Dependence of Effective Diffusivity on Temperature

The diffusivities of the species in water are calculated by a linear approximation to the Nernst equation for electrolytes (18) and by the Wilke-Chang equation for dissolved gases (32):

For electrolytes:

$$D_{eT} = D_{eT_0} (1 + 0.04 T) \quad (17)$$

For dissolved gases:

$$D_{eT} = 5.06 \times 10^{-9} \frac{T}{\mu V_b^{0.6}} \quad (18)$$

Equation (17) is used for NH_4^+ , NO_3^- , and HCO_3^- , and Equation (18) is used for O_2 and H_2CO_3 .

REACTOR MODEL

The hydraulics and flow regime in a fluidized bed are modeled in terms of (1) the residence time distribution of the fluid in the reactor, (2) the statics and dynamics of fluidization, and (3) the effect of the reactor flow regime on external mass transfer of reacting species to the biofilm surface.

Residence Time Distribution

Residence time distribution studies of the pilot plant reactor have shown that nearly all axial dispersion occurs at the bed inlet and is caused by the inlet distribution system (26). The residence time distribution created by the entrance effects move virtually without change through the upper portion of the column, which seems to closely approximate plug flow. This is shown by Figure 11.

If the reactor is long compared to its diameter, mixing at the inlet region becomes insignificant and the overall regime approaches plug flow. If the bed is short compared to its diameter a mixing model is required to describe reactor residence time distribution.

The model chosen for this study is the backmix-reactors-in-series model. Residence time distribution information in this model is contained in the empirical parameter n . The reactor is imagined to be a sequence of n equal volume completely stirred reactors. The concentration calculated in reactor $(i-1)$, which is also its effluent concentration, is the inlet concentration to reactor (i) . As n becomes large, the model approaches the plug flow model.

The reactor is shown schematically in Figure 12. The model is given by:

$$V_i \frac{dC_{ai}}{dt} = F(C_{a,i-1} - C_{ai}) \pm V_i R(C_{ai}) \quad (19)$$

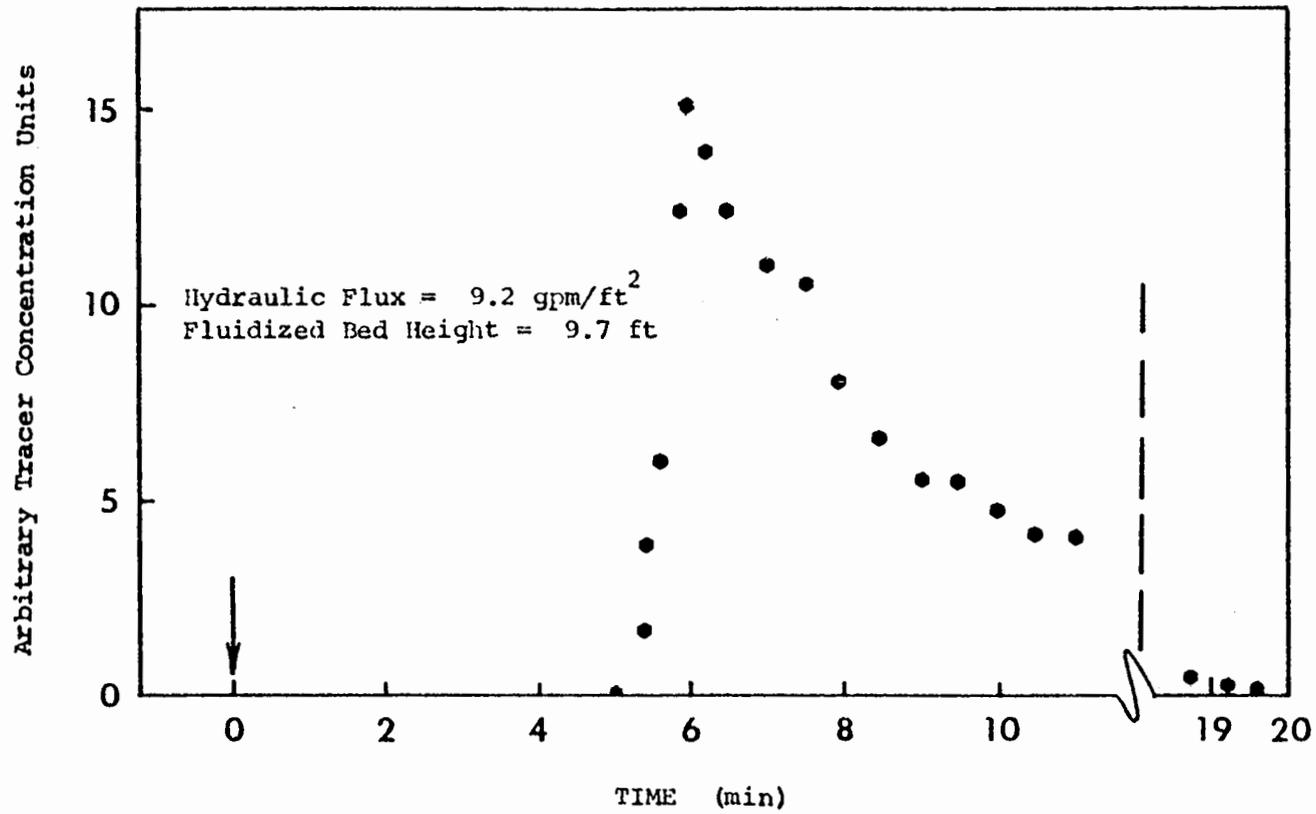


Figure 11 Typical Residence Time Distribution for the Fluidized Bed Pilot Reactor

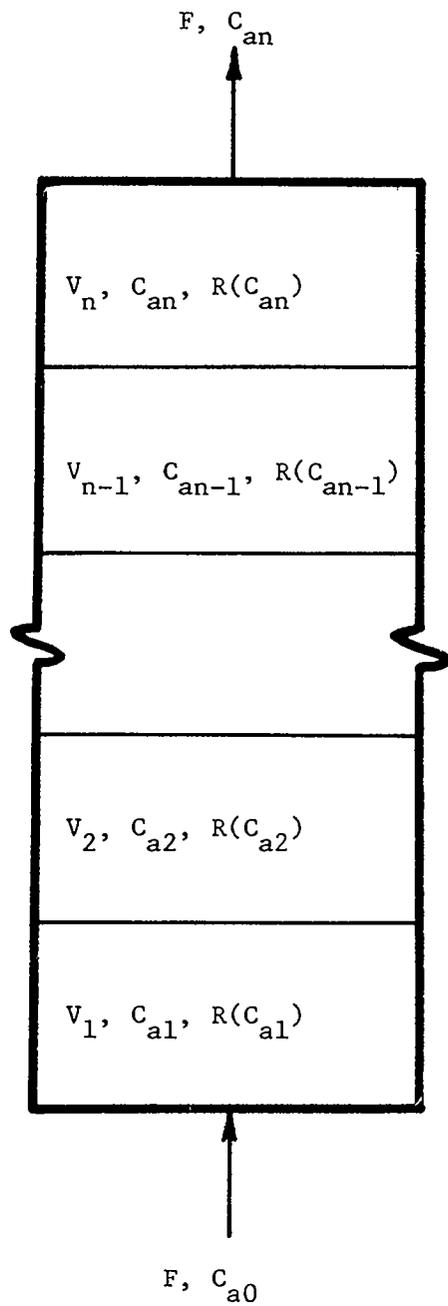


Figure 12 Schematic for Backmix-Reactors-in-Series Flow Model

for $i = 1, n$. In this equation, V_i is the volume of segment i (equal to the total reactor volume divided by n), F is the volumetric flow rate, and C_{ai} is the concentration of species A in segment i . The function $R(C_{ai})$ is the sum of the surface fluxes of species A into the bioparticles.

Fluidization

A satisfactory model describing steady state fluidization is by Fair and Hatch (10) which uses the Carmen-Kozeny equation (16) for the headloss in a granular bed of constant particle size modified for particles of equivalent sphere diameter, d_i , and sphericity, ψ :

$$\left(\frac{h_f}{l}\right)_i = \frac{kU\mu}{g\rho} \frac{(1 - f_i)^2}{f_i^3} \left(\frac{6}{\psi d_i}\right)^2 \quad (20)$$

If the bed is fluidized, the frictional headloss through the bed is equal to the buoyant weight of the suspended particles:

$$\left(\frac{h_f}{l_e}\right)_i = \frac{(\rho_s - \rho)}{\rho} (1 - f_i) \quad (21)$$

Substitution of (21) into (20) and rearrangement gives:

$$f_i^3 - (1 - f_i) \frac{kU\mu}{g(\rho_s - \rho)} \left(\frac{6}{\psi d_i}\right)^2 = 0 \quad (22)$$

This can be solved using Newton's method, for f_i for each particle size.

The total depth of the expanded bed is:

$$L = L_o (1 - f_o) \sum_i \frac{P_i}{(1 - f_i)} \quad (23)$$

The particle diameter and particle density in a fixed film system must be modified to account for the presence of the film. The carrier particle diameter d_i is increased by addition of twice the biofilm thickness.

$$d_i = d_{ic} + 2\delta_i \quad (24)$$

The modified particle density is calculated from the volumes and densities of the clean particle and the biofilm:

$$\rho_p = \rho_{pc} \left(\frac{d_{ic}}{d_i}\right)^3 + \rho_f \left(1 - \left(\frac{d_{ic}}{d_i}\right)^3\right) \quad (25)$$

where ρ_f is the wet density of the biofilm. It is calculated from the dry film density (ρ_d) and the moisture content (P):

$$\rho_f = \frac{\rho_d}{1 - P} \quad (26)$$

The net effective particle diameter and density values are used in the fluidization calculation, Equation (22).

Fluidization Dynamics

Fluidization dynamics have been modeled by Slis et al. (24) using the Richardson-Zaki (23) model for steady state fluidization:

$$U_{ss} = v_s f_{ss}^m \quad (27)$$

and give the transient change in bed depth as:

$$\frac{dL}{dt} = U_\infty - v_s f_L^m \quad (28)$$

where U_∞ is the superficial velocity after a step increase in flow and f_L is the instantaneous porosity at the top of the bed.

Fan et al. (11) linearized this model at an initial value by Taylor series expansion to obtain:

$$\frac{dL}{dt} = \frac{1}{\theta} (L_\infty - L) \quad L = L_{ss} \text{ at } t = 0 \quad (29)$$

The time constant θ is calculated from the initial conditions by:

$$\theta = \frac{L_{ss} f_{ss}}{U_{ss}^m (1 - f_{ss})} \quad (30)$$

Since the exponent m is not well defined for beds of non-uniform particle size, θ should be determined experimentally for a particular sand to account for its particle size distribution.

The integrated form of the linear model, Equation (29):

$$L = L_{ss} \exp(-t/\theta) + L_{\infty}(1 - \exp(-t/\theta)) \quad (31)$$

is satisfactory for step increases in flow.

The response to a step decrease in flow differs mechanistically from that for the step increase. The response is zero order (24) and:

$$\frac{dL}{dt} = -K \quad (32)$$

with $L = L_{ss}$ at $t = 0$, and $L = L_{\infty}$, at $t = \infty$. The rate constant K equals the change in superficial velocity resulting from the step decrease in flow rate:

$$K = U_{\infty} - U_{ss} \quad (33)$$

Equations (29) through (33) have adequately described the fluidization dynamics in the pilot plant fluidized bed (26).

External Mass Transfer

External mass transfer is the transport of species through the stagnant liquid boundary layer of the bioparticle to the particle surface. In some systems this limits the overall rate of removal from the bulk solution. In a fluidized bed, there is a vigorous general circulation of particles and high local velocities and this limitation should not be important, as the following analysis indicates.

Under ideal conditions, the flux of a species through a stagnant fluid layer is:

$$N_s = k_1 (S_b - S_s) \quad (34)$$

where N_s is the flux of S , k_1 is the overall mass transfer coefficient, and S_b and S_s are the concentrations of the species in the bulk fluid and at the solid surface.

The overall mass transfer coefficient, k_1 , incorporates diffusive and convective mass transfer processes. The complex

flow patterns in a porous bed make it impossible to estimate k_1 from first principles. Many correlations have been developed to estimate k_1 from the system geometry and flow conditions (4). For fluidized beds, Gupta and Thodos (12) report the following correlation for mass transfer from spheres into liquids:

$$k_1 = \frac{G Sc^{-2/3}}{\rho f} 0.01 + \frac{0.863}{R_e^{0.58} - 0.483} \quad (35)$$

which holds for $R_e > 1$. In this equation, G is the mass velocity of the fluid, R_e is the particle Reynolds number, and Sc is the Schmidt number.

Typical values in a fluidized bed are $R_e = 3.27$, $Sc = 670$, $\bar{f} = 0.5$ and $G = 0.364$ ($\text{g cm}^{-2}\text{sec}^{-1}$) (26). Using these values in equation (35) gives an estimate of $k_1 = 0.0055$ cm sec^{-1} as a typical value for the overall external mass transfer coefficient in a fluidized bed.

The effect of external mass transfer on the removal rate depends on the intrinsic reaction rate (i.e., the rate which would be obtained in the absence of mass transfer resistance). For a single limiting substrate, the Monod rate expression is:

$$-r = \frac{q_m S}{K_s + S} \quad (36)$$

If the system is at steady state, $N_s = -r$, and:

$$k_1 (S_b - S) = \frac{q_m S}{K_s + S} \quad (37)$$

Introducing the dimensionless variables $\bar{S} = S/S_b$ and $\bar{K} = K_s/S_b$ converts Equation (37) to:

$$\frac{q_m}{k_1 S_b} \frac{\bar{S}}{\bar{K} + \bar{S}} - (1 - \bar{S}) = 0 \quad (38)$$

The dimensionless ratio $q_m/k_1 S_b$ is known as the Daimkohler number (D_a). This important parameter describes the relative magnitudes of species flux due to reaction and

flux due to external mass transfer across the stagnant layer. For $D_a > 1$, the mass transfer rate is said to be limiting; for $D_a < 1$, the reaction rate is limiting.

Typical values for q_m and S_b in a fluidized bed for nitrification are $0.064 \text{ mg/cm}^2\text{hr}$ and 5 mg/l-N . Using the value of k_1 calculated in Equation (35) gives $D_a = 0.646$, which is less than 1 so the error in ignoring external resistance will be small. Therefore external mass transfer resistance is not included in the model.

MODEL CALIBRATION AND VERIFICATION

Steady state calibration runs at different hydraulic fluxes were done five times during the pilot plant study. Influent ammonium concentration, pH, alkalinity, and temperature were uncontrolled, but variation was negligible during a run.

Two important unmeasured variables in the model were the biofilm thickness and the active mass of nitrifying bacteria. The model was calibrated to other data to estimate these variables. For each run the height of fluidization was calculated by adjusting the biofilm thickness until the calculated and observed height agreed, after which, the dry density of the nitrifying biomass was adjusted so the model predicted the observed ammonium concentrations at different heights in the reactor. The high influent dissolved oxygen concentration could not be measured precisely; it was estimated from the effluent concentration and reaction stoichiometry. Alkalinity, pH, and nitrate measurements were used to verify the calibration. The results of one of these calibration runs are shown in Figure 13. The good agreement of the data and the calculations are typical of the calibrations obtained.

The fitted values for film thickness, nitrifier density, and influent dissolved oxygen are given in Table III. Biofilm thickness is seen to increase with time; observation of the system confirms this calculated trend. This is expected since bed solids were not controlled. The slight decrease in δ between runs 3 and 4 is a result of a hydraulic flux increase between the runs. The turbulence caused visible loss of biomass from the bed, most likely those solids that were only loosely attached to the sand. The subsequent increase in film density needed to fit the concentration data tends to support this. This reduction in film thickness would increase the density from 10 to 14 mg/cm^3 at most if

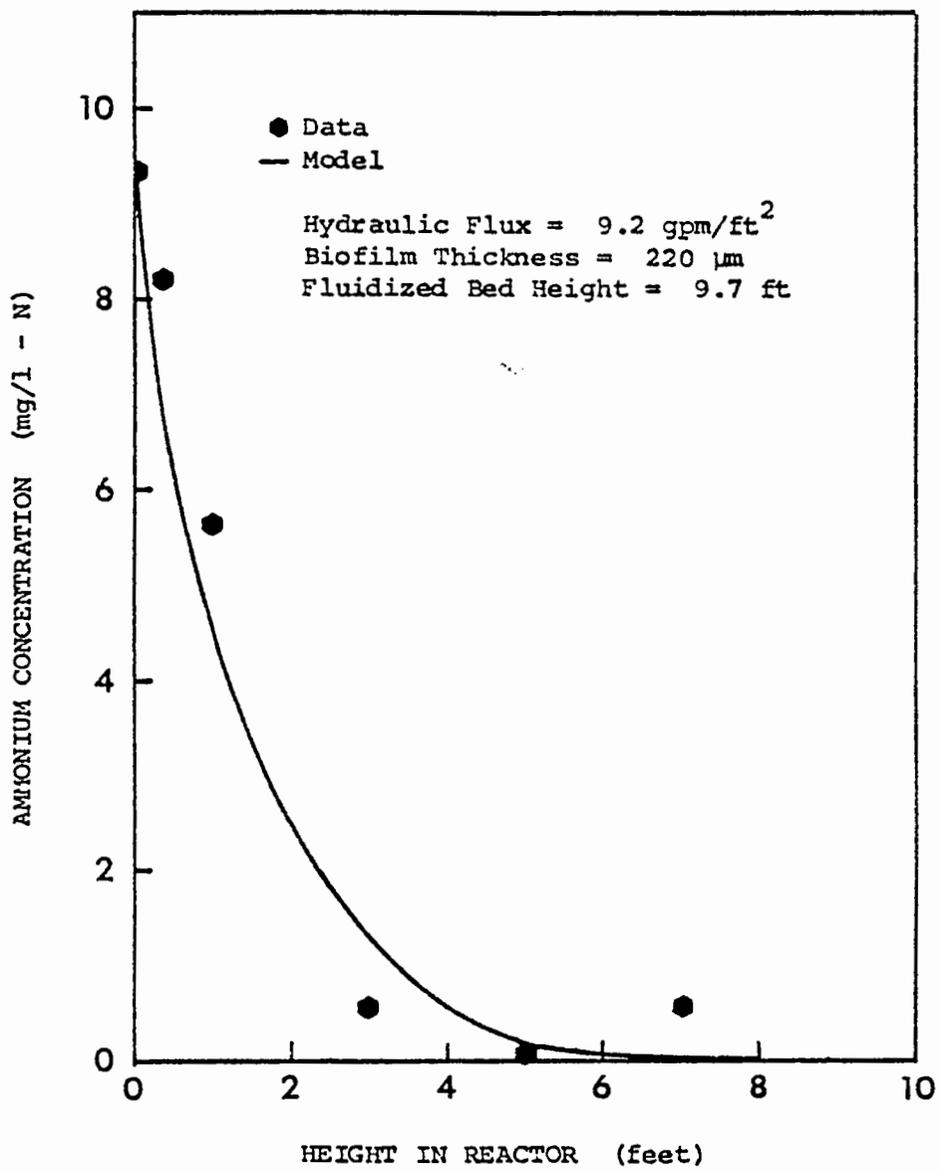


Figure 13 Reactor Ammonium Profile -- Dec. 18, 1980
Model Calibration Results

negligible mass was associated with the lost solids. For these purposes, however, unknown changes occurring during the flow increase are accommodated in ρ_d . The film thickness in run 5 is larger than in run 4 while the density has decreased, which may indicate that the solids that accounted for the increase in film thickness are less dense than in run 4. The model assumes a uniform film thickness and density throughout the bed. Deviations from this assumption may cause some of the apparent changes in thickness and density.

Table III. Values of Calibration Parameters Used in Model Fitting

Run No.	Date	Film Thickness (cm)	Film Density (mg/cm ³)	Influent Dissolved Oxygen (mg/l)
1	11/12	0.003	13	55
2	11/24	0.014	15	61
3	12/13	0.022*	10	75
4	12/13	0.018**	20	75
5	12/18	0.022	17	60

* Before step increase in flow.

** After step increase in flow.

The dissolved oxygen concentrations estimated in Table III are higher than the 50-55 mg/l cited earlier in this paper. Performance of the oxygen feed system was variable and the influent concentrations of 50-55 mg/l are estimates of the average over the course of the study. The saturation concentration of dissolved oxygen at the operating pressure of 30 psig is about 135 mg/l at 20°C so it is possible for the influent dissolved oxygen to be as high as 75 mg/l at times. Also, the higher influent D.O. estimates are near the end of the study where temperatures were low and the saturation concentration of oxygen is higher than earlier in the study. The values obtained, although not typical of 'normal' operation are reasonable in light of the uncertainties involved.

The model gives good agreement with the pilot plant ammonium data with just three adjustable parameters. The trends in these parameters with time can be explained qualitatively. Steady state verification is provided by the

measurement of pH, HCO_3^- , and NO_3^- . Results for run 5 are plotted in Figures 14 and 15. Alkalinity was not measured for this run. The close agreement of calculated and observed values provides verification for the steady state model.

DYNAMIC SIMULATION

After verification of the steady state model, the unsteady state model was used to predict the response of the process to the dynamic forcings discussed previously. The simulations are plotted in Figures 5, 6 and 8 along with the experimental results for each run. No additional fitting parameters were used in their calculation.

Step Forcing of Ammonium

Figure 5 shows a good prediction of the shape and location of the response curve. However, the measured steady state ammonium concentration before the step is lower than the prediction by about 2 mg/l-N. This calibration run was done after the step change was made. Apparently the influent D.O. was lower after the step change than before it was made. The oxygen flow rate probably decreased when the step was made. This would have reduced the ammonium removal capacity. The influent dissolved oxygen concentration was not accurately known due to limitations in the analytical test used. Effluent dissolved oxygen decreased from 2.5 mg/l to less than 0.5 mg/l during the experiment. Considering these uncertainties, agreement between the model and the observed response is encouraging.

Step Forcing of Flow

In this test, the hydraulic flux to the bed was doubled; the bed expanded and the empty bed detention time decreased from 12.7 to 7.3 minutes. The measured and predicted responses are plotted in Figure 6. The flow increase caused turbulence which sheared flocculent biomass from the particles in the bed. The resulting steady-state fluidized bed height was lower than it would have been without the biofilm loss. This hydraulic change also increased the apparent biofilm density because the solids lost were of relatively low density.

The model predicts negligible ammonium breakthrough into the effluent. Exact numerical comparison with the data is

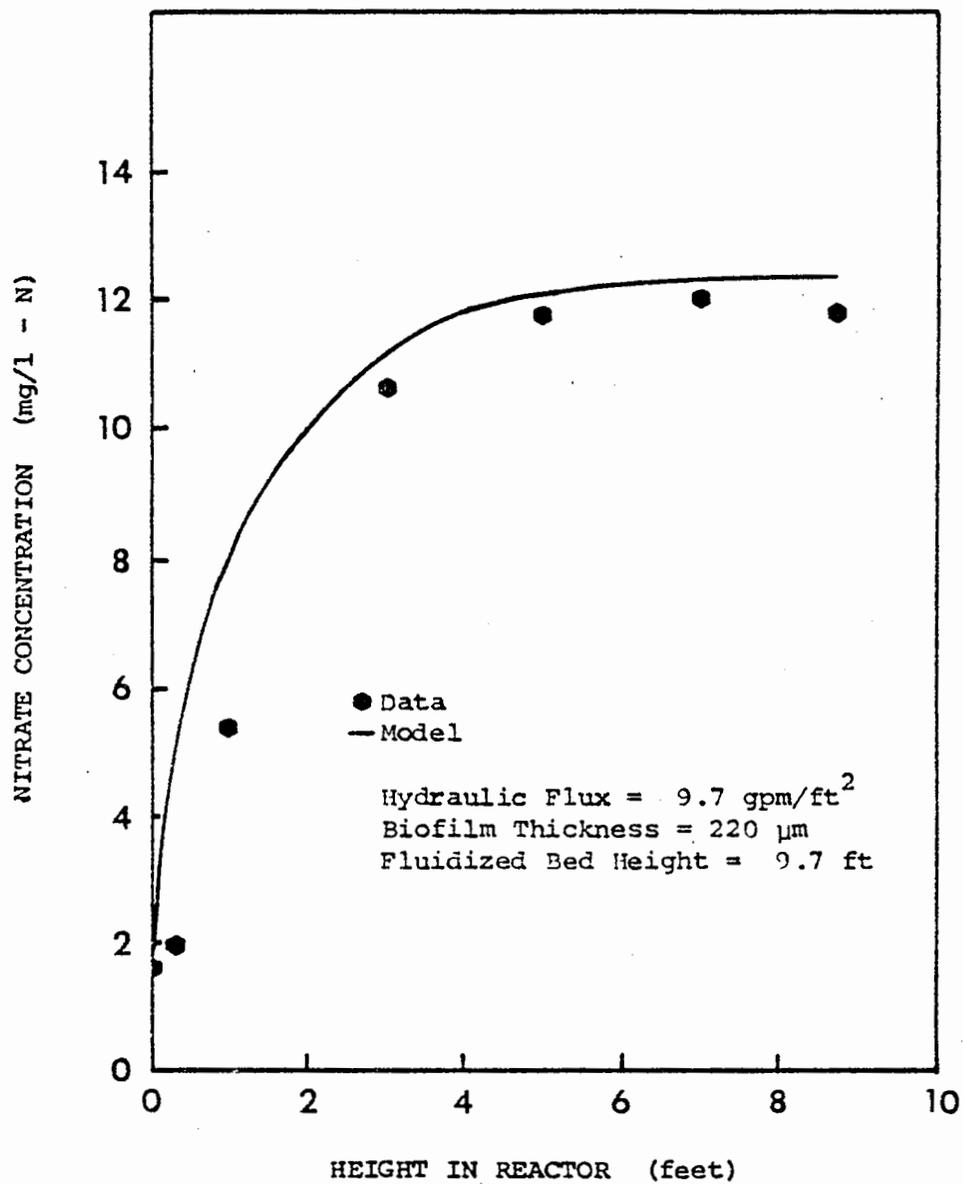


Figure 14 Reactor Nitrate Profile -- Dec. 18, 1980
Model Verification Results

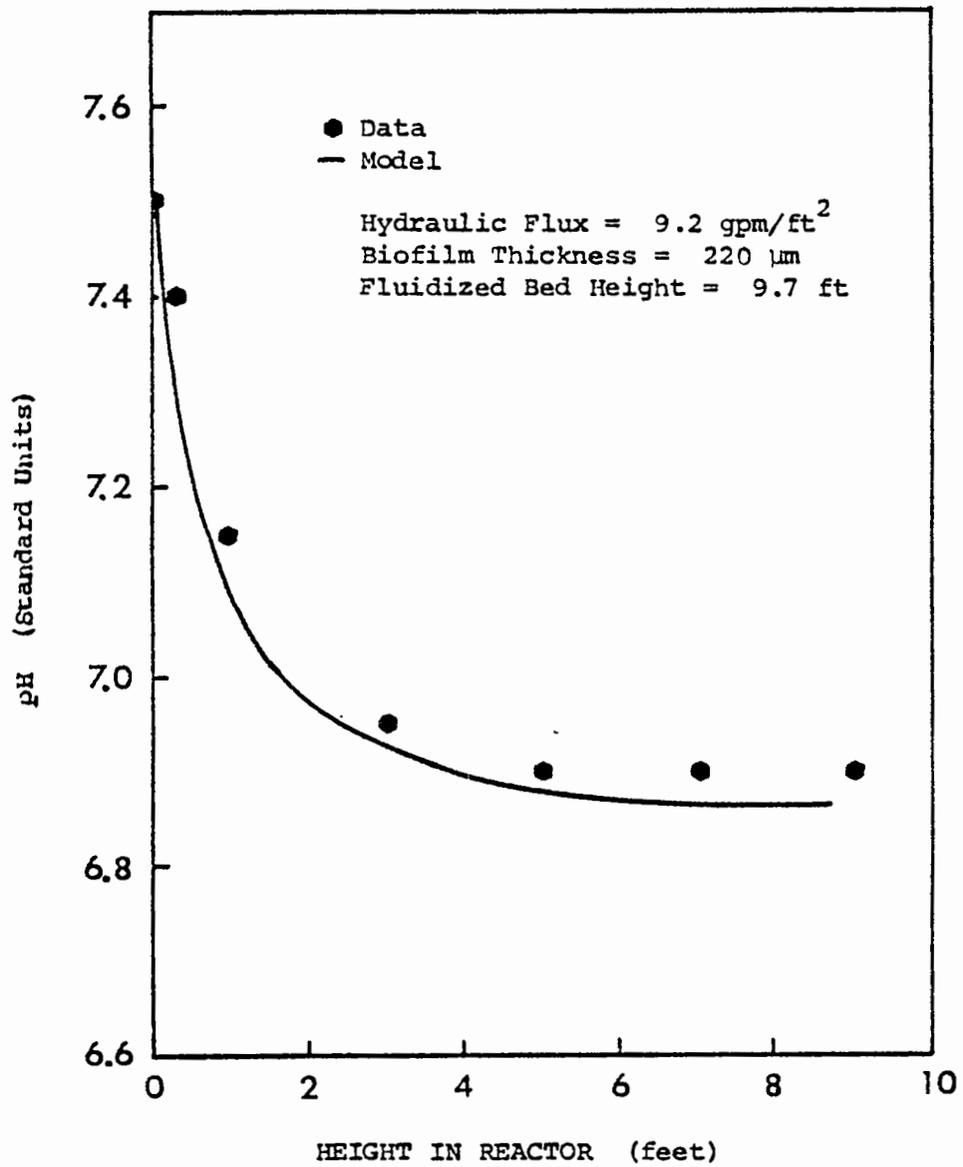


Figure 15 Reactor pH Profile -- Dec. 18, 1980
Model Verification Results

difficult since the observed concentrations approach the limit of sensitivity of the analytical test. The flow increase also caused unsteady mixing patterns in the bed which could account for the random variations in the response that cannot be predicted by the model. The prediction is very good considering the low concentrations involved.

Impulse Forcing of Ammonium and Nitrite

The model does not calculate the nitrite response so only the ammonium results will be discussed. The measured response of the process and the model prediction are shown in Figure 8, for an impulse of 185 mg $\text{NH}_4^+\text{-N}$. The model accurately predicted the rising and falling legs of the curve and the location and height of the peak. The simulated response curve was integrated to estimate that 54% of the ammonium added was removed; integrating the data gives 53% removed.

Summary

The calibration runs indicated that the kinetic model, which includes rate limitations from several substrates and intraparticle diffusion, adequately describes the rate of substrate flux into the bioparticles. Alkalinity, pH, and nitrate data give good verification of the steady state prediction. The values of parameters used to fit the experimental data seem typical and their trends with time can be qualitatively explained. The parameters obtained in the steady state runs were used to successfully simulate the dynamic experiments. The results of these dynamic experiments lend general validity to the theoretical approach and provide a basis for further work to eliminate the uncertainty incorporated in the fitting parameters.

CONCLUSIONS

The main objective of this research was to study the behavior of the fluidized bed nitrification process under steady-state and time-varying conditions. A mathematical model has been developed to describe the system under these conditions. As a result of this research it may be concluded that:

- 1) Secondary effluent at Madison Nine Springs treatment plant can be successfully treated for ammonium

removal by the fluidized bed process. Empty bed hydraulic detention times in the range of 10 to 15 min. are sufficient.

- 2) The bed volatile solids were high, which confirms previous studies. The short detention times observed are also typical.
- 3) Failure to supply sufficient oxygen to the system was the major factor that limited ammonium removal capacity when the efficiency was less than essentially 100 percent.
- 4) The fluidized bed absorbed a doubling of hydraulic flux with negligible deterioration of effluent quality. It was able to absorb over 50% of an impulse forcing of 185 mg ammonia; more could be removed with a larger oxygen supply.
- 5) The mathematical model adequately predicted the steady state and dynamic response of the system.
- 6) The multiple substrate model of the reaction kinetics makes the model effective over a range of operating conditions.
- 7) The most uncertainty in the model is in the film thickness and nitrifier density. More accurate measurements of influent dissolved oxygen are also needed. Research efforts should be directed toward more precise measurement of these three variables.

NOMENCLATURE

b	decay coefficient (T^{-1})
C	species concentration (ML^{-3})
C_{ai}	species A concentration in stage i (ML^{-3})
C_b	species concentration in bulk solution (ML^{-3})
d_i	particle diameter of fraction i (L)
d_{ic}	diameter of carrier particle i (L)
Da	Daimkohler number (\dagger)
D_e	effective diffusivity in biofilm (L^2T^{-1})
D_{eT}	effective diffusivity at temperature T (L^2T^{-1})
D_{eT_0}	effective diffusivity at temperature T_0 (L^2T^{-1})
f_i	porosity of bed due to particles of d_i (\dagger)

f_o	porosity of unexpanded bed (\dagger)
f_L	porosity at top of fluidized bed (\dagger)
f_{ss}	porosity of bed at steady state (\dagger)
F	volumetric liquid flow rate (L^3T^{-1})
g	gravitational constant (LT^{-2})
G	mass velocity ($ML^{-2}T^{-1}$)
h_f	headloss (L)
k	permeability coefficient (\dagger)
k_l	mass transfer coefficient (LT^{-1})
K	rate constant for fluidization dynamics (LT^{-1})
\bar{K}	dimensionless parameter K_s/S_b
K_a	stability constant for carbonate equilibrium ($molesL^{-3}$)
K_s	saturation coefficient (ML^{-3})
l	bed depth (L)
l_e	expanded bed depth (L)
L	bed height (L)
L_o	unexpanded bed height (L)
L_{ss}	steady state bed height (L)
L_∞	steady state bed height after change (L)
m	parameter in Richardson-Zaki equation (\dagger)
n	number of stages in reactor model (\dagger)
p_i	mass fraction of particles of d_i (\dagger)
P	moisture content (\dagger)
q_m	maximum substrate removal rate ($ML^{-3}T^{-1}$)
r	radial distance from particle center (L)
R_f	bioparticle radius (L)
R_p	carrier particle radius (L)
$R(C)$	reaction rate of C ($ML^{-3}T^{-1}$)
Re	Reynolds number (\dagger)
S	substrate concentration (ML^{-3})

S_b	bulk substrate concentration (ML^{-3})
S_s	surface substrate concentration (ML^{-3})
\bar{S}	dimensionless substrate concentration S/S_b
Sc	Schmidt number (\dagger)
t	time (T)
T	temperature (degrees)
U	superficial fluid velocity (LT^{-1})
U_{ss}	steady state superficial fluid velocity (LT^{-1})
U_∞	steady state superficial fluid velocity after change (LT^{-1})
v_s	terminal settling velocity of particle (LT^{-1})
V_b	partial molar volume of gas ($L^3 \text{mole}^{-1}$)
V_i	volume of stage i (L^3)
X^*	biomass density (ML^{-3})
Y^*	yield coefficient (M cells/M substrate)
δ	biofilm thickness (L)
θ	time constant for fluidization dynamics (T^{-1})
μ	fluid viscosity ($ML^{-1}T^{-1}$)
$\hat{\mu}^*$	maximum specific growth rate (T^{-1})
$\hat{\mu}_T$	maximum specific growth rate at temperature T (T^{-1})
$\hat{\mu}_{max}$	maximum specific growth rate at pH 8.3-9 (T^{-1})
ρ	fluid density (ML^{-3})
ρ_d	dry biofilm density (ML^{-3})
ρ_f	wet biofilm density (ML^{-3})
ρ_p	bioparticle density (ML^{-3})
ρ_{pc}	carrier particle density (ML^{-3})
ρ_s	general solid particle density (ML^{-3})
ψ	particle sphericity (\dagger)

* subscripts and superscripts for these terms are: n (NH_4^+), o (O_2), b (BOD), h (HCO_3^-), h₂ (H_2CO_3). † dimensionless.

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HYDRODYNAMICS OF FLUIDIZED BED REACTORS FOR WASTEWATER TREATMENT

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INTRODUCTION

Fluidized bed reactors offer significant advantages for aerobic and anaerobic treatment of wastewater, compared to activated sludge and fixed bed biological treatment systems. The major advantages are stated in several recent articles and books (2, 9, 10, 20, 22). Most common are: reduced reactor volume, due to the large surface area for microbial growth provided by the support particles and the elimination of clogging that occurs in fixed bed systems.

The significance of hydrodynamics; in particular, the effects of hydrodynamic instability are often underestimated when fluidized bed reactors are modelled. This may cause difficulties when processes developed at the pilot plant level are translated into large scale treatment installations. In the concluding paragraph of their book Gelperin, N. I., Aishtein, V. G., and Kwasha, V. B. (6) state, "At the present time, only the division of a full-size reactor by vertical baffles into sections of the same dimensions as the size of the test model, fed with the same amount of fluidizing agent in each section insures the same conversion of

matter as that obtained in the model." Significant progress has been made, since this book was published in 1967. For example, improvements in water distribution systems and the introduction of calming sections above the distributor have resulted in minimizing instabilities, resulting from maldistribution of the flow of liquid into the bed. Also mathematical models describing the transition from a stable to an unstable regime of fluidization (5, 18), have been developed, however, these models neither describe a mechanistic cause for the loss of stability, nor do they provide a means of restoring the system to a stable regime.

For liquid-solid systems hydrodynamic instability may occur at both low and high velocities of flow. At low velocities instability may be exhibited in the form of channeling. The flow will pass through only a fraction of the bed, resulting in unutilized capacity. At high velocities large circulation patterns of the fluidized material may be observed, a phenomenon commonly referred to as "boiling". Both of these forms of instability result in the reduction of process efficiency.

For practical purposes, the conditions which lead to the loss of stability should be determined and a technical means, which insure a stable regime of fluidization should be developed.

In a first approach, evaluation of the stability can be performed based on the deviation of system parameters from an idealized steady state model, which describes the expansion relationship of the bed, the relationship between superficial velocity of the liquid and the average porosity of the bed for particles of given properties (size, shape, and density) at constant temperature. The model developed by Mints, D. M. (11, 16, 17) describes this ideal fluidization of solid particles by a liquid for stable regimes of fluidization. The model is presented in this paper and supported with our experimental data and data published by Wilhelm, R. H. and Kwauk, M. (24). Deviations from ideality are presented graphically and evaluated qualitatively. Finally, the effects of a new type of fixed packing, intended to increase the stability of fluidized beds is discussed.

IDEAL MODEL OF LIQUID-SOLID FLUIDIZED BEDS

A number of phenomenological models describing the expansion behavior of fluidized beds are presented in the literature (1,3,5,6,11,13,16,17,19,21). Mints' model can be

used to adequately describe this relationship and is preferred to the other models, since it is developed on a better theoretical basis, a relationship between Reynolds number of the flow through porous media and the resistance of the media. It incorporates the properties of the particles and the liquid.

The model was developed to describe the hydrodynamics of backwashing for filters and was verified experimentally by Mints for relatively large and dense materials, such as sand, gravel, and steel balls. Extension of this model to smaller particles and materials with a low density relative to water was one of the objectives of the present work. In accordance with the model, the upward flow of fluid through the suspended particles is considered to be a special case of filtration. The values of resistance coefficient, C_d , and Reynolds number, Re , are related in the following manner:

$$ReC_d = ReC_o + m \quad (1)$$

where C_o and m are dimensionless coefficients.
In the general case:

$$C_d = \frac{Pl}{H\rho_1u^2} \quad (2)$$

$$Re = \frac{\rho_1 u \ell}{\mu} \quad (3)$$

where $u = V/\epsilon =$ true velocity of flow in the porous space,
 $\ell = \epsilon/w =$ hydraulic radius of the porous space,
 $V =$ superficial (empty column) velocity),
 $\epsilon =$ porosity of the bed,
 $P = (\rho_2 - \rho_1)g(1 - \epsilon) =$ hydraulic resistance of the bed,
 $\rho_2 =$ density of the solid,
 $\rho_1 =$ density of the fluid,
 $\mu =$ viscosity of the fluid.

Substitution of the values of u , ℓ , and P into expressions (2) and (3) yields:

$$C_d = \frac{\rho_2 - \rho_1}{\rho_1} g \frac{\epsilon^3 d}{6v^2} \quad (4)$$

$$Re = \frac{\rho_1 v d}{6\mu(1 - \epsilon)} \quad (5)$$

For a single particle with a terminal settling velocity, θ , the values of resistance coefficient, C_p , and Reynolds number, Re_p , are

$$C_p = \frac{\rho_2 - \rho_1}{\rho_1} g \frac{\pi d}{6\theta^2} \quad (6)$$

$$Re_p = \frac{\rho_1 \theta d}{\mu} \quad (7)$$

Comparisons of (4) and (6) and (5) and (7) yield:

$$C_d = C_p \frac{\epsilon^3}{\beta^2 \pi} \quad (8)$$

$$Re = Re_p \frac{\beta}{6(1-\epsilon)} \quad (9)$$

where $\beta = V/\theta$ = ratio of the superficial velocity of flow through suspended layer to the terminal settling velocity of discrete particle.

Combining expressions (1), (8), and (9) results in the following equation:

$$C_p \frac{\epsilon^3}{\beta^2 \pi} = \frac{6m(1-\epsilon)}{Re_p \beta} + C_o \quad (10)$$

In the case of settling of single particle $V=\theta$, $\beta=1$, and $m=1$, equation (10) degenerates into the relationship $C_o=C_p/\pi$. Substituting this relationship into (10) and solving for the value of $\beta>0$ yields:

$$\beta = \frac{V}{\theta} = - \frac{3\pi m(1-\epsilon)}{Re_p C_p} + \sqrt{\left[\frac{3\pi m(1-\epsilon)}{Re_p C_p} \right]^2 + \epsilon^3} \quad (11)$$

or

$$\beta = - m^1(1-\epsilon) + \sqrt{[m^1(1-\epsilon)]^2 + \epsilon^3} \quad (12)$$

where

$$m^1 = \frac{3\pi m}{Re_p C_p} = \text{dimensionless hydrodynamic characteristic of the particles.}$$

Expressing the porosity, ϵ , in relationship (12) through the concentration of the suspension, $C = \rho_2 g (1-\epsilon)$, or the volumetric concentration, $C_v = (1-\epsilon)$, and multiplying

both sides of the equation by θC_v , an expression describing Kynch's curve, flux versus concentration, is obtained. This equation contains the terminal velocity of a particle, θ , and an empirical parameter, m^1 . In order to facilitate the determination of these parameters, equation (10) can be linearized by substituting the relationship $C_d = C_p/\pi$ into (10), multiplying it by πV , and dividing by $C_p(1-\epsilon)\theta^2$. This results in the following expression:

$$\frac{\epsilon^3}{(1-\epsilon)V} = \left[\frac{1}{\theta^2}\right] \frac{V}{(1-\epsilon)} + \left[\frac{2m^1}{\theta}\right] \quad (13)$$

This expression relates the measurable parameters, V and ϵ . Plotting the experimental values in $y = \epsilon^3/[(1-\epsilon)V]$ versus $x = V/(1-\epsilon)$ produces a straight line with a slope, $1/\theta^2$, and the segment cut on the ordinate $2m^*/\theta$. The plot, in accordance with equation (13), is especially convenient when particle characteristics, such as diameter, shape factor, etc. are unknown.

VERIFICATION OF THE IDEAL MODEL

It was stated in the previous section that Mints verified the model for large and relatively dense materials. However for fluidized bed biological reactors it is desirable to use relatively small and light particles, because of the advantages of the greater specific surface area and relative ease of fluidization afforded by these particles. For this reason experiments were performed with "small" sand, granular activated carbon, and crushed black walnut shells to verify Mints' model for these materials. Additional data was obtained from the literature, Wilhelm, R. H. and Kwauk, M. (24). The experimentally determined physical properties of the particles is presented in Table 1. The specifics of the experimental apparatus and procedures are described in the following subsection.

Experimental Installations and Procedures

Two rectangular plexiglass columns with internal dimensions 60 x 10 x 1.27 cm³ were used by the authors. The columns had a thin third dimension to enhance visual observations of the fluidized bed through backlighting. A schematic of the small experimental column is presented in Figure 1.

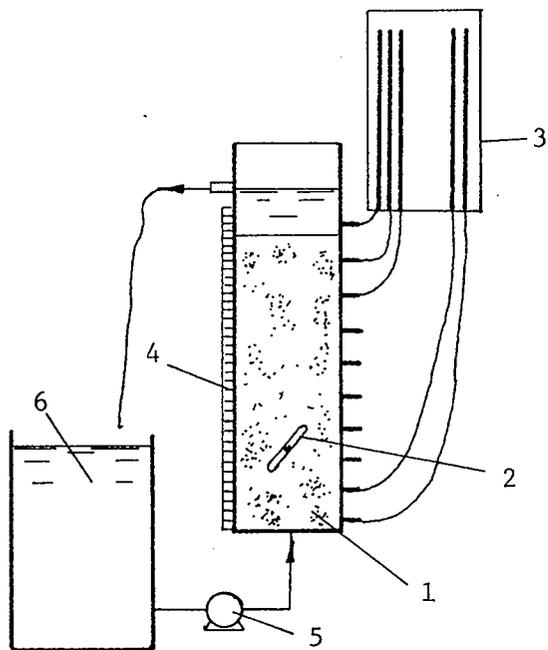


Figure 1. Schematic of experimental installation.

- 1 - flat column fragment, 2-plate,
- 3 - piezometers, 4-millimetre scale,
- 5 - circulation pump, 6-water tank.

Copper tubing with evenly spaced holes directed downward was located near the base of the column for water distribution. Water was collected in a hose located at the top of the column, leading to a 50 gal. tank for recycle by pumping. The flow rate was controlled by the use of a by-pass. Piezometers were located at 5 cm intervals along the length of the column.

The large column had sixteen brass fittings, located near the base of the column. These fittings were attached to the exterior (eight on the front and eight on the back), directed downward at a 45° angle. This configuration permitted the plugging of any number of holes to vary the flow distribution across the width of the column. A flow splitter was constructed to distribute the flow to the inlet fittings. Water entered the splitter through a 3/4 in. diameter pipe and exited through brass fittings, linked to those at the base of the column by plastic tubing. Water was discharged over a weir at the top of the column and collected in a plexiglass box attached to the back of the column. This box was used for flow measurement by clamping the discharge lines and measuring the time required to fill a known volume.

Material used in the tests was sieved with the fraction retained on the sieve corresponding to the desired grain-size removed. This fraction was re-sieved until only a small percentage passed the sieve. The material was then placed in a column through which flow was passed, causing the material to fluidize. Hydraulic classification would occur. The finest material was wasted over the weir by increasing the flow. The heaviest material was removed by siphoning from the base of the column. The density of the material was determined by the use of a pycnometer. Water absorbing materials were saturated overnight in a vacuum flask. They were spread out on a towel and the surface of the particles was wiped dry. The equivalent diameter, d , was determined by the method of counting and weighing particles. The terminal settling velocity of the particles was determined by dropping particles and measuring the time to travel an assigned distance.

Fluidization experiments were performed to obtain data for verification of Mints' model. The porosity of the media, ϵ_0 , was determined by placing material of known weight in a column, compacting it to a dense state by tapping the column with a rubber hammer, and measuring the volume of the bed. This value is related to the average porosity of the fluidized bed by the relationships: $C_v = C_{v0} (h_0/H)$ and $\epsilon = (1-C_v)$, where h_0 is the height of the compacted bed and H is the

height of the fluidized bed. Flow of water was introduced into the bottom of the column, causing the bed to fluidize. The bed was allowed to reach a steady level. The flow rate was determined by measuring the time necessary to collect a specific volume of water. The velocity, V , determined from flow measurements, was checked by measuring the subsidence rate of the upper surface of the bed after the flow into the column was stopped. Several tests were performed to determine the expansion rate of the bed.

Wilhelm, R. H. and Kwouk, M. performed their tests in 3 inch and 6 inch diameter columns. For each run the flow rate was gradually increased in a series of steps. Measurements of the flow rate, headloss, and porosity were taken at each step. The experimental apparatus and procedures are discussed in detail in the article (24).

Analysis of Experimental Data

The experimental data was quantitatively analyzed by use of the linearized forms of Mints' model, Equations (10) and (12), and by the use of Kynch's plot, solid flux versus concentration. Plots of the data are presented in Figures 2 through 19. The plots show good linearization of the data; however, systematic deviations occur at low values of Reynolds number, Re and a random scatter occurs at high Reynolds numbers. The systematic deviations can be seen mainly for fine sand ($d=0.04$ cm) in Figure 6. For Reynolds numbers less than 1, the data plots a concave curve. For Reynolds numbers greater than 1, the fit of experimental points is almost perfect. This behavior is also exhibited by small glass beads ($d=0.05$ cm) in Figure 12, where the data for Reynolds numbers less than 1 appear obviously deviate from linearity adjacent to the vertical axis. Due to this consistent deviation for both sets of data, these points were not included in the computations to determine the parameters of Mints' model. Experimental points for low values of Reynolds numbers greater than 1 also depart from linearity. For this regime liquid channels through the bed, which may be expanded but is not fluidized. An empirical correlation to determine the minimum fluidization was developed by Wen, C. Y. and Yu, Y. H., c. f. Cleasby, J. and Fan, K. (3):

$$Ga = \frac{d^3 \rho_1 (\rho_2 - \rho_1) g}{\mu^2} \quad (14)$$

where

Ga = Galileo number

$$\text{Re}_{mf} = \frac{\rho_1 V_{mf}}{\mu} = [(33.7)^2 - 0.0408 \text{ Ga}]^{0.5} - 33.7 \quad (15)$$

Note that the linear parameter included in Reynolds number, Re_{mf} , is different than that in the Reynolds number, Re , defined in Mints' model. This correlation was used as a criterion for the elimination of points corresponding to the channeling bed, for velocities less than the minimum fluidization velocity.

Computations were performed to determine the parameters C_0 , and m from a linear regression of Equation (10), using the particle properties d , ρ_2 , and Θ , and values of ϵ and V from fluidization experiments. The results of the linear regression were used to compute the empirical parameter $m' = 3 m / \text{Re}_p C_p$.

A shape factor, or sphericity term could have been included in the presentation of Mints' model, in the form, Ψd , where Ψ is the sphericity and d is the particle diameter in Equations (6) and (7). This term would compensate for the irregular shape of non-spherical particles. Although it is not explicitly stated in the equations, the sphericity is implicitly contained in the coefficient m' , in Equation (12), which is the most convenient form for determining the parameter, describing the expansion behavior of the bed.

For verification of the model, this term would directly affect the value of the diameter of the particle, d , and the value of the settling velocity, Θ . However, since the value of the settling velocity is determined experimentally, rather than by computations from particle properties; the diameter of the particles is the only parameter which would be directly affected. Therefore, neglecting the sphericity term and assuming spherical particles introduced small errors in the calculations, in view of the accuracy with which the particle diameter can be determined experimentally. Additionally the sand, granular activated carbon, and ground black walnut shells used by the authors appeared visually to be "rounded". Therefore, for computational purposes the particles were assumed to be spherical.

Another set of computations independent on particle properties was performed, involving a linear regression in accordance with Equation (12). The values of velocity and porosity determined during the fluidization experiments were

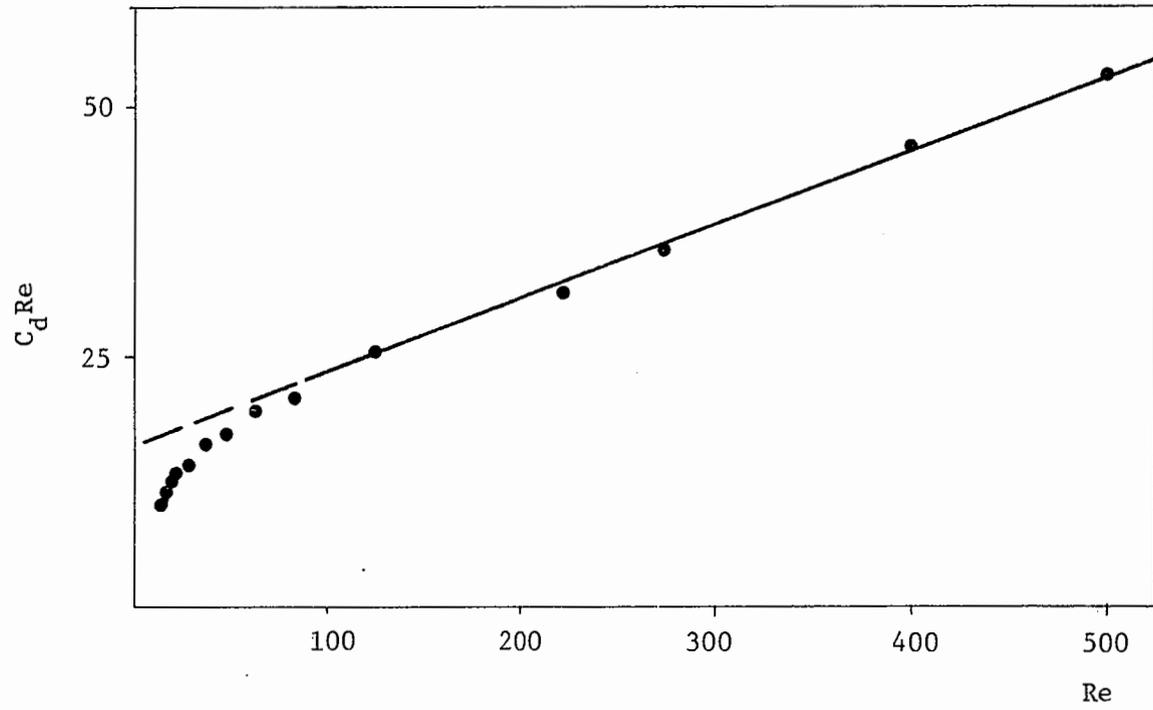


Figure 2. Relationship between resistance coefficient and Reynolds number.

(Lead Shot)

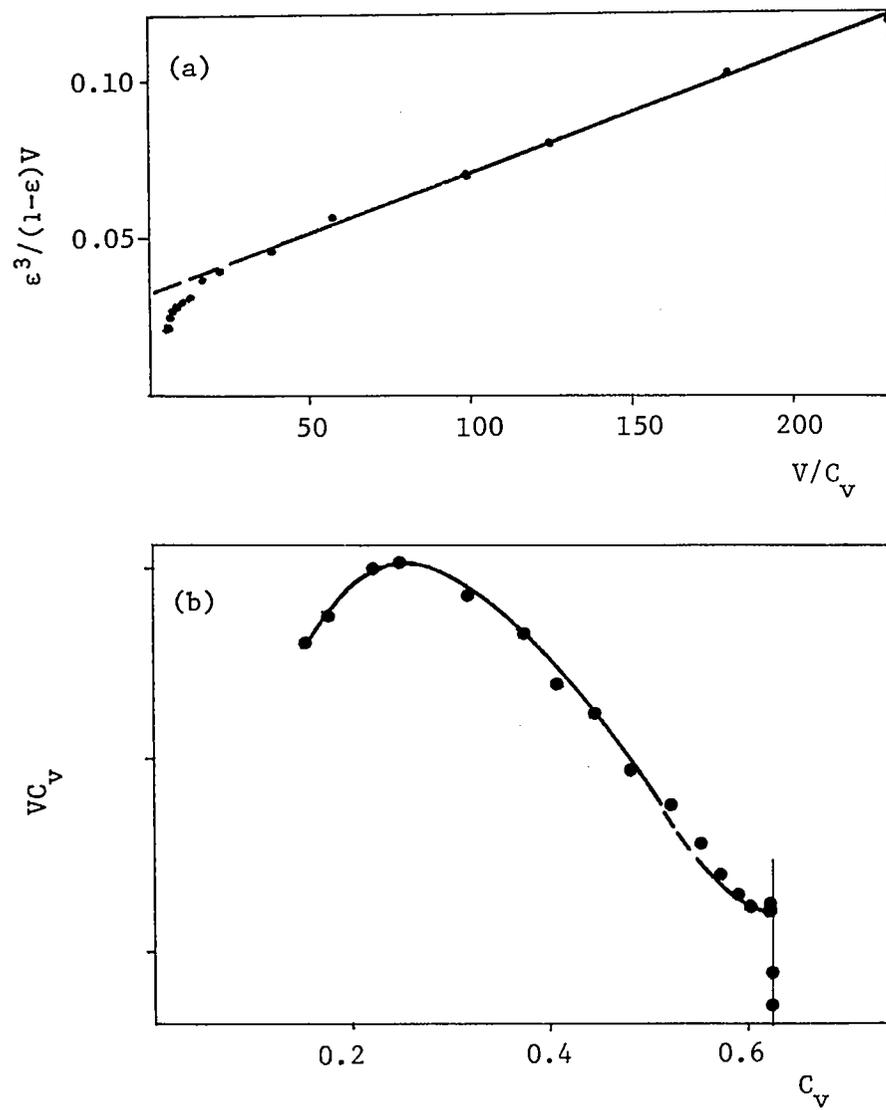


Figure 3. Mints linearized plot (a) and Kynch curve (b).

(Lead shot)

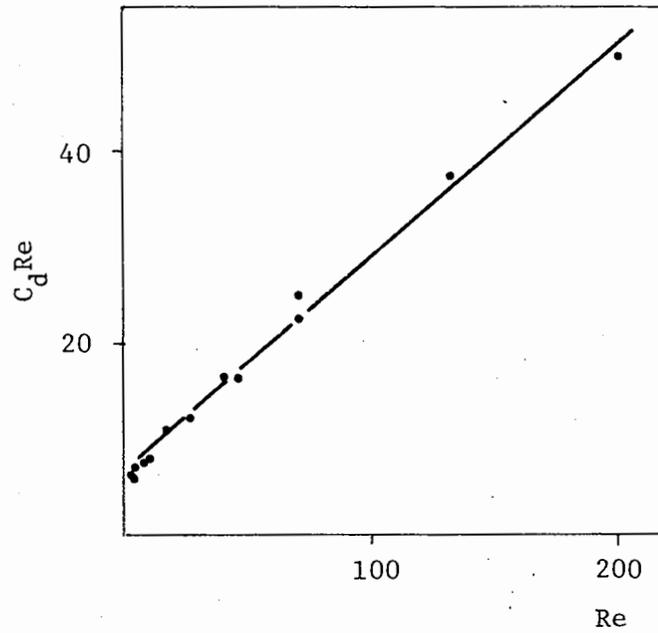


Figure 4. Relationship between resistance coefficient and Reynolds number.

(Sea sand, $d = 0.10$ cm)

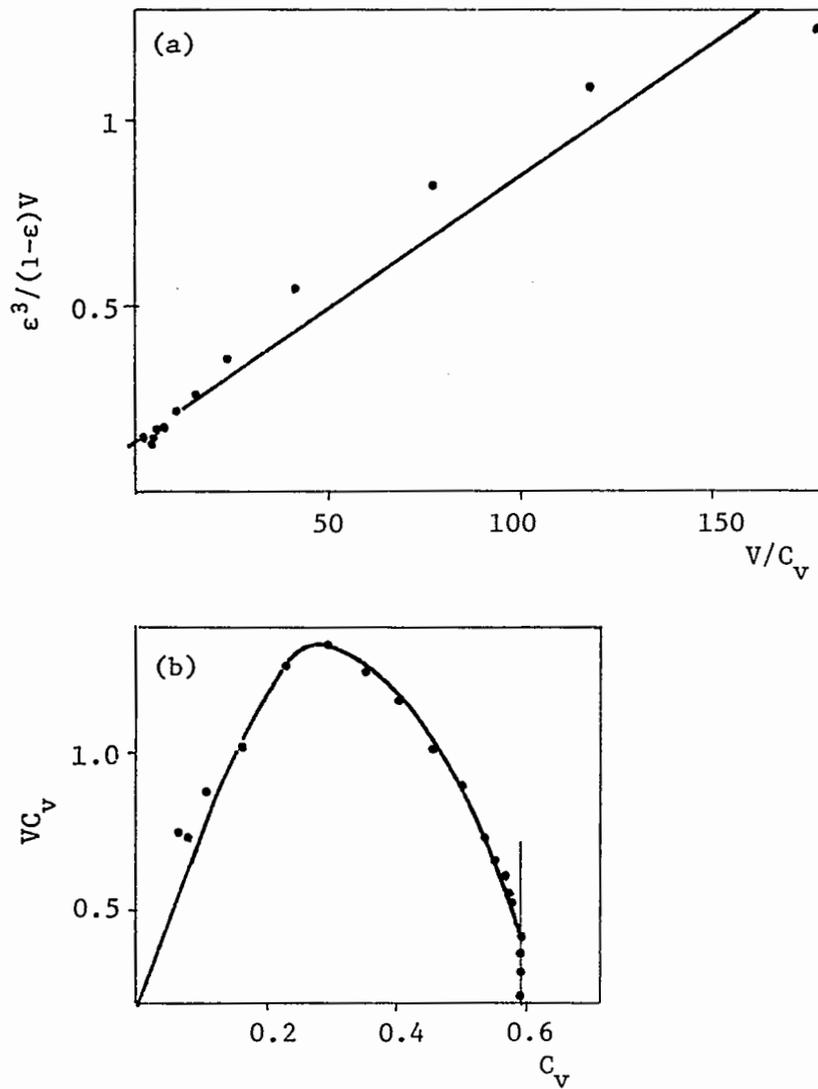


Figure 5. Linearized Mints plot (a) and Kynch Curve (b).

(Sea sand, $d = 0.10$ cm)

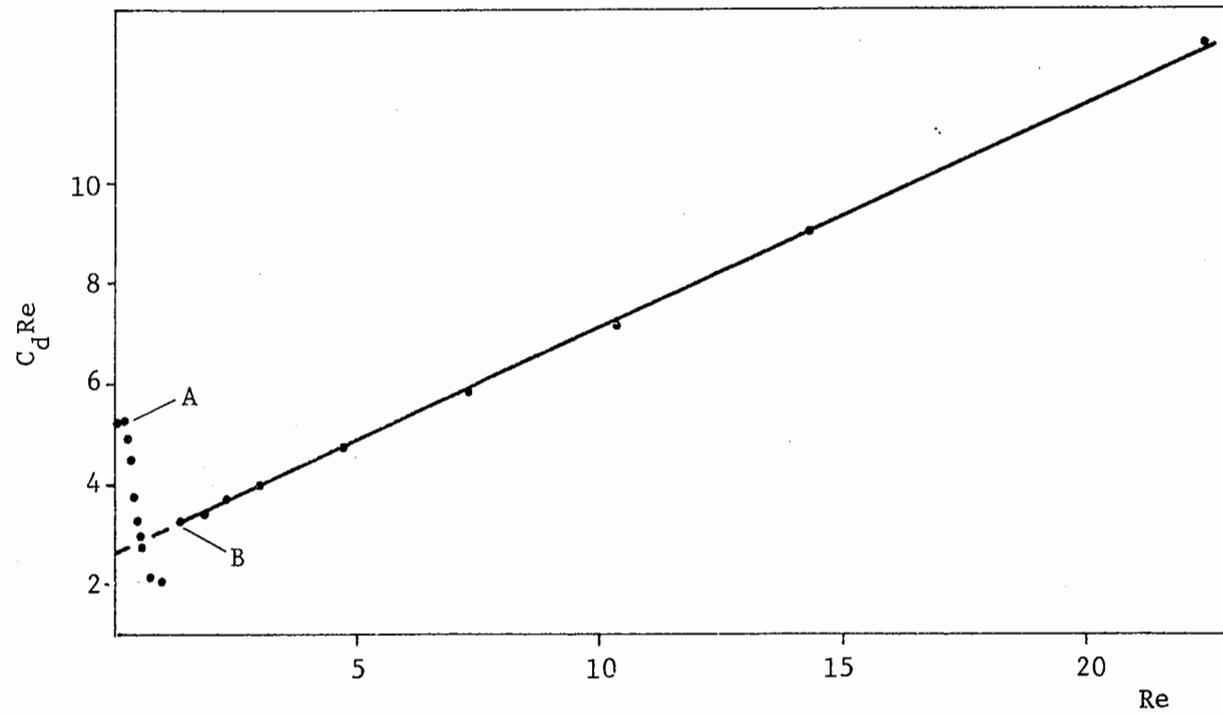


Figure 6. Relationship between resistance coefficient and Reynolds number.

(Sea sand, $d = 0.037$ cm)

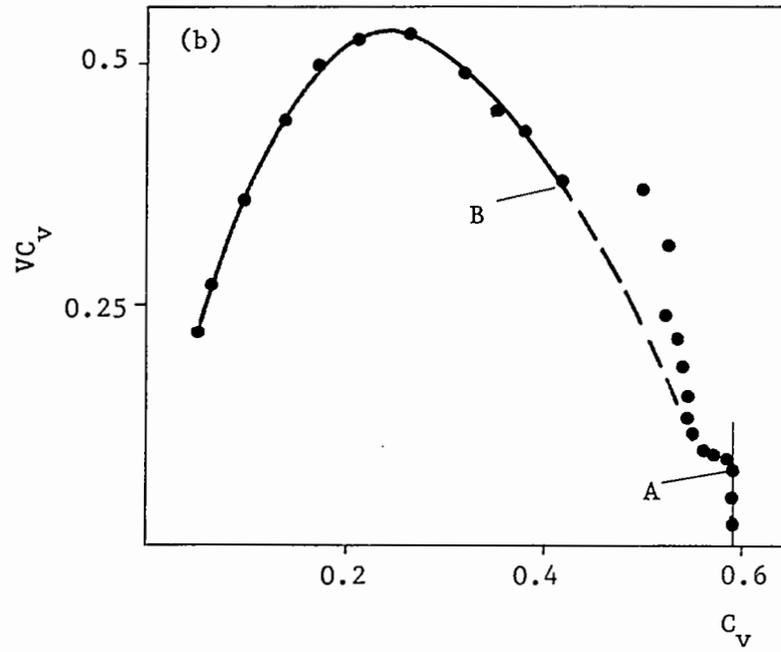
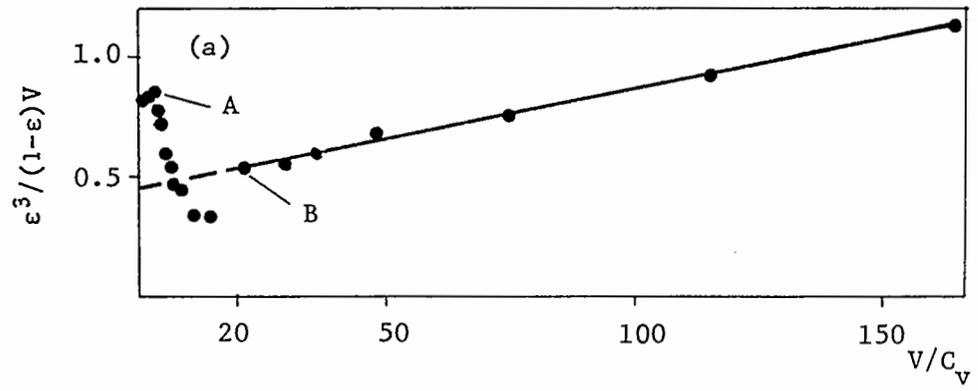


Figure 7. Mints linearized plot (a) and Kynch curve (b).

(Sea sand, $d = 0.037$ cm)

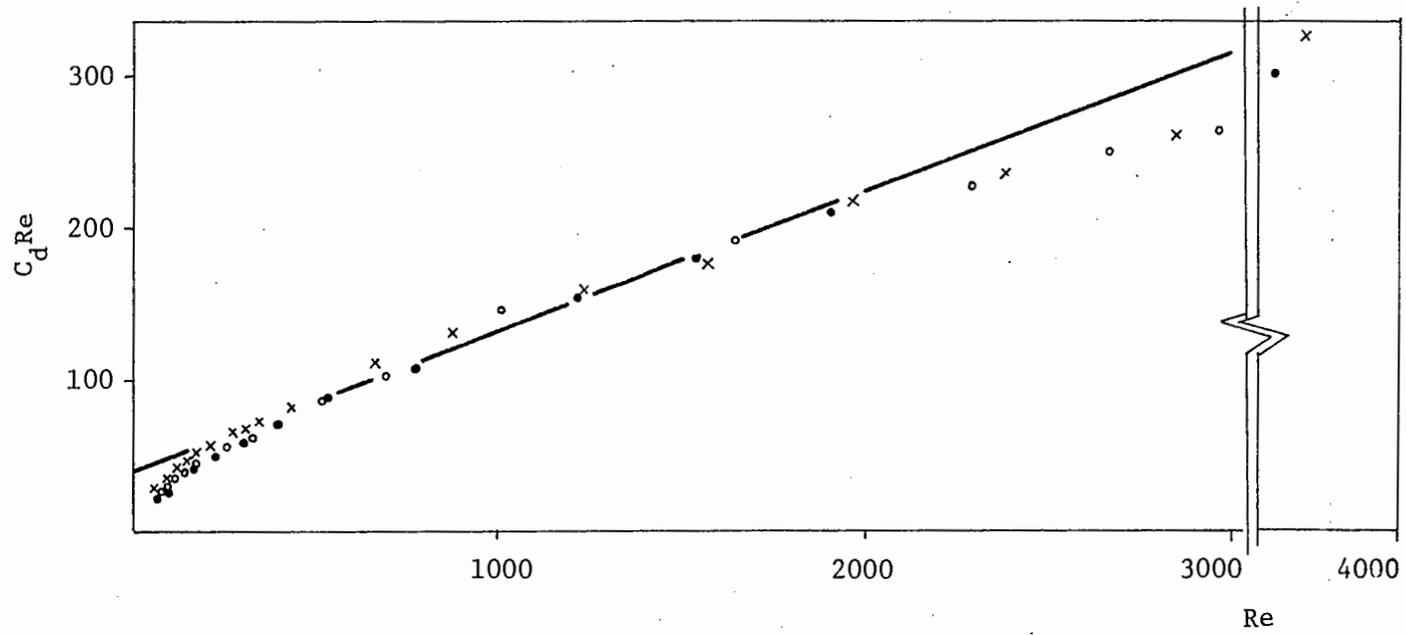


Figure 8. Relationship between resistance coefficient and Reynolds number.

(Glass beads, $d = 0.521$ cm,

$\times - h_o = 6.2$ cm, $o - h_o = 12$ cm, $\bullet - h_o = 19.5$ cm)

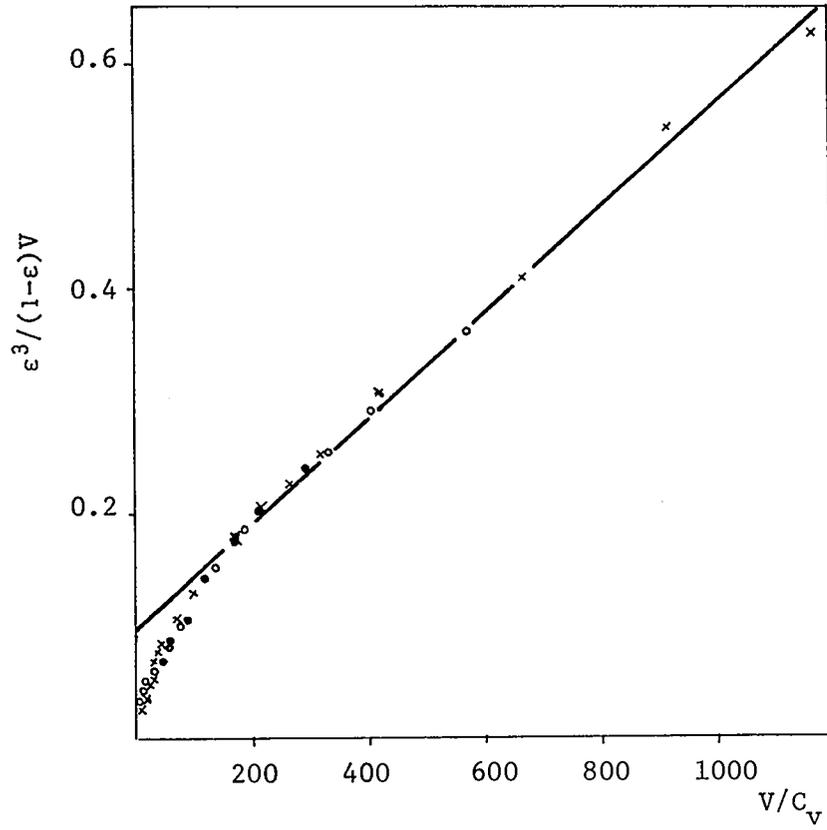


Figure 9. Mints linearized plots

(Glass beads, $d = 0.521$ cm,

$\times - h_0 = 6.2$ cm, $o - h_0 = 12$ cm,

$\bullet - h_0 = 19.5$ cm)

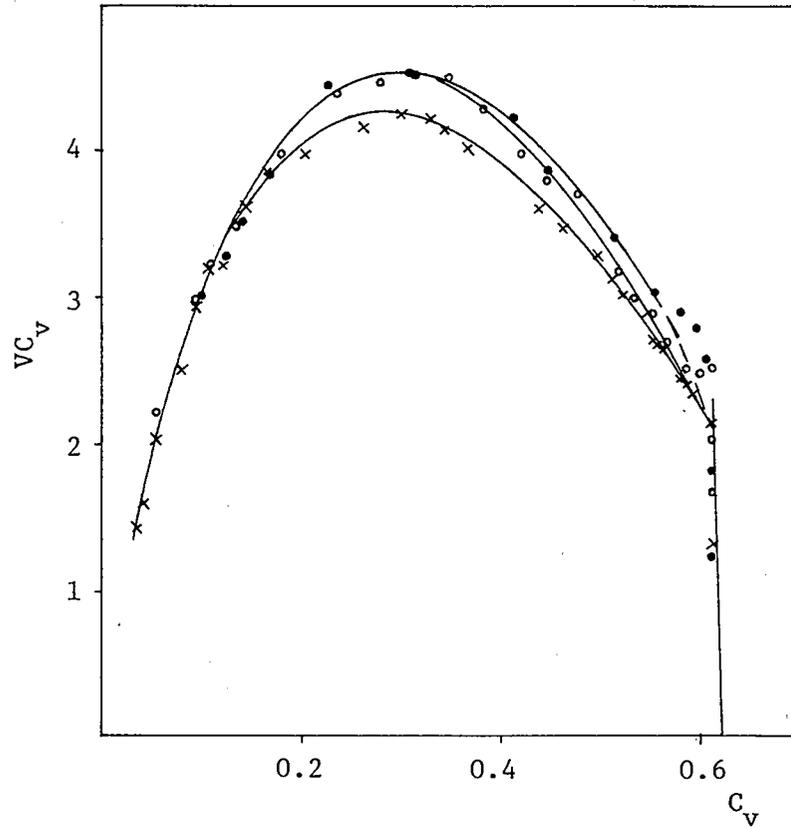


Figure 10. Kynch curves

(Glass beads, $d = 0.521\text{cm}$

$x-h_0 = 6.2\text{ cm}$, $0-h_0 = 12\text{ cm}$,

$\bullet-h_0 = 19.5\text{ cm}$).

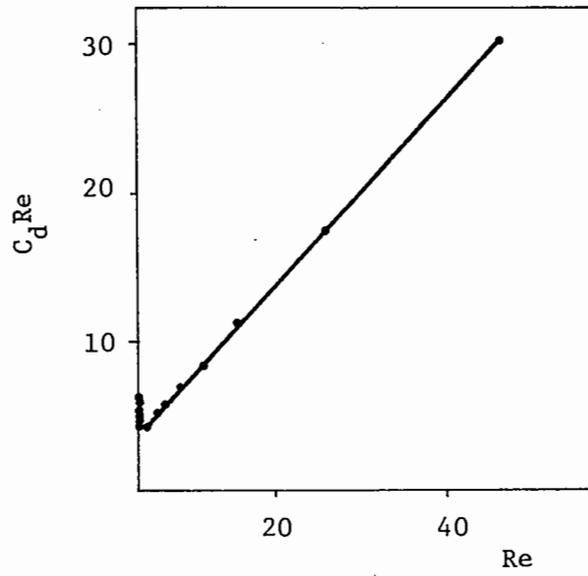


Figure 11. Relationship between resistance coefficient and Reynolds number.

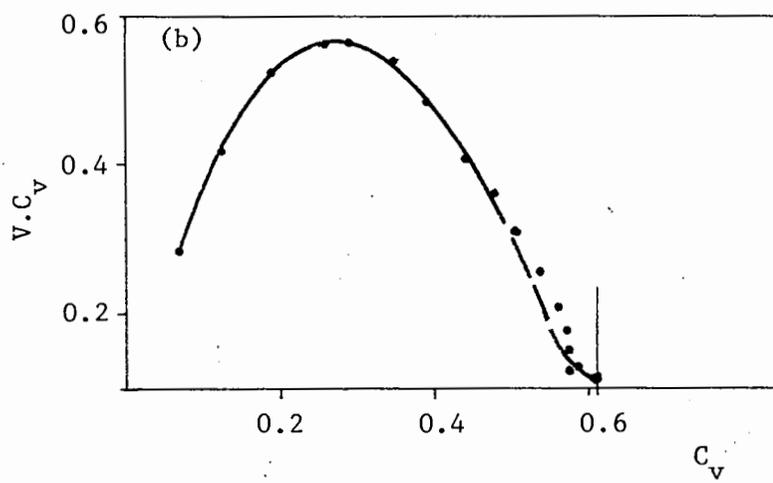
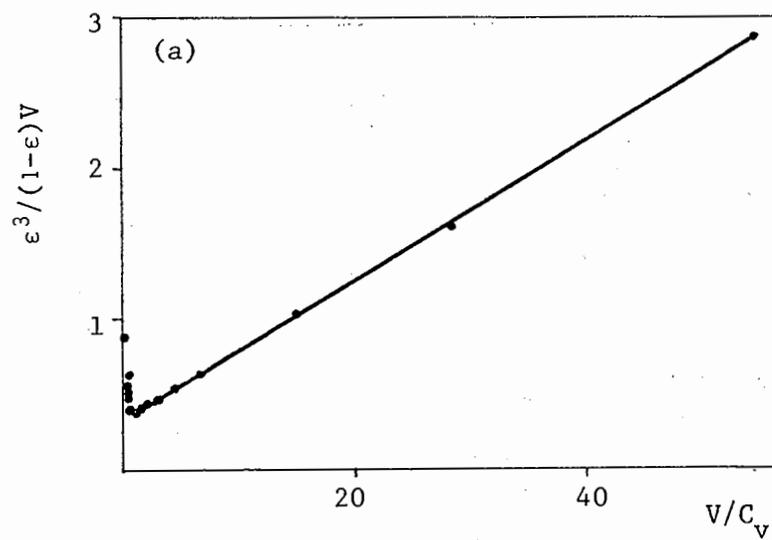


Figure 12. Mints linearized plot (a)
and Kynch curve (b).
(Glass beads, $d = 0.051$ cm)

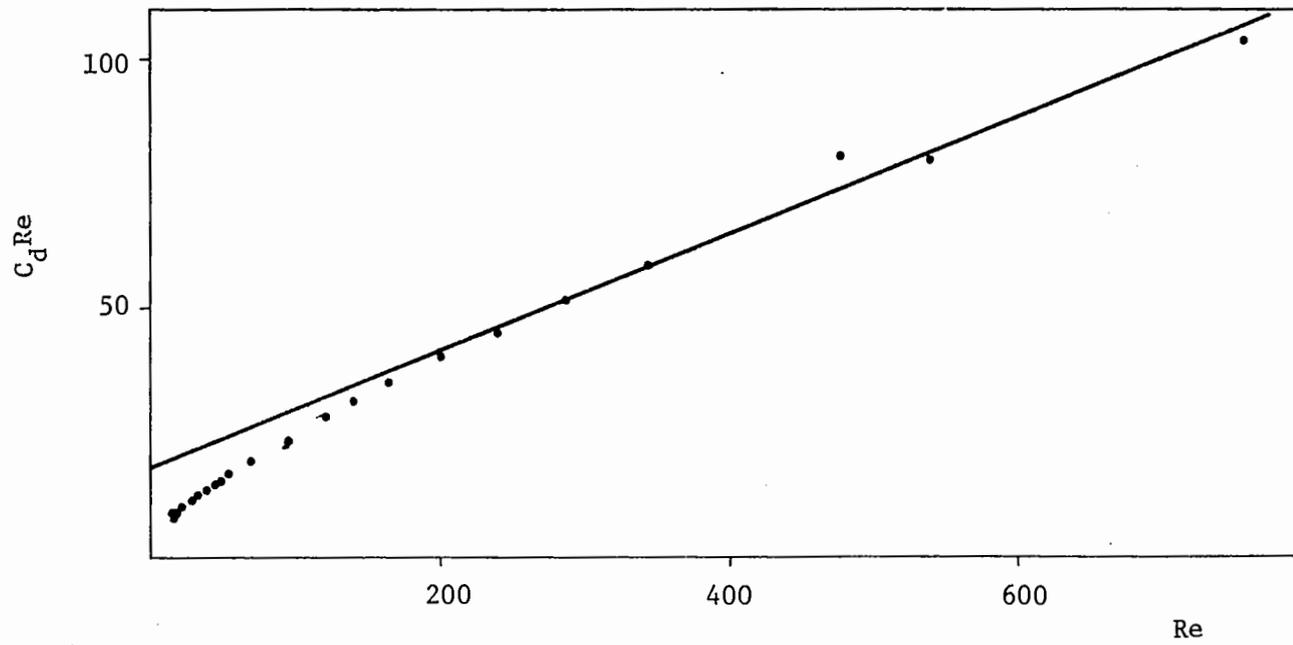


Figure 13. Relationship between resistance coefficient and Reynolds number.

(Sacony beads, $d = 0.328$ cm)

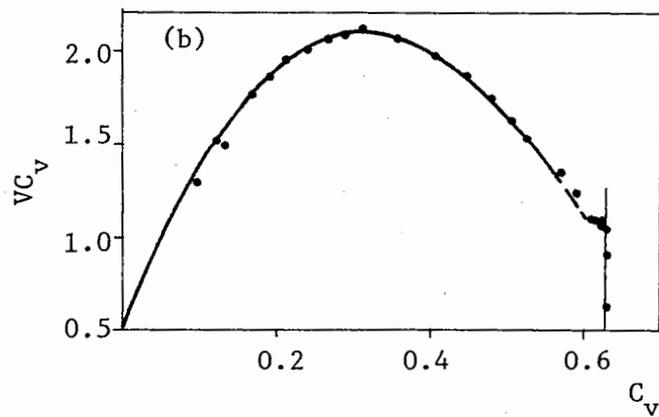
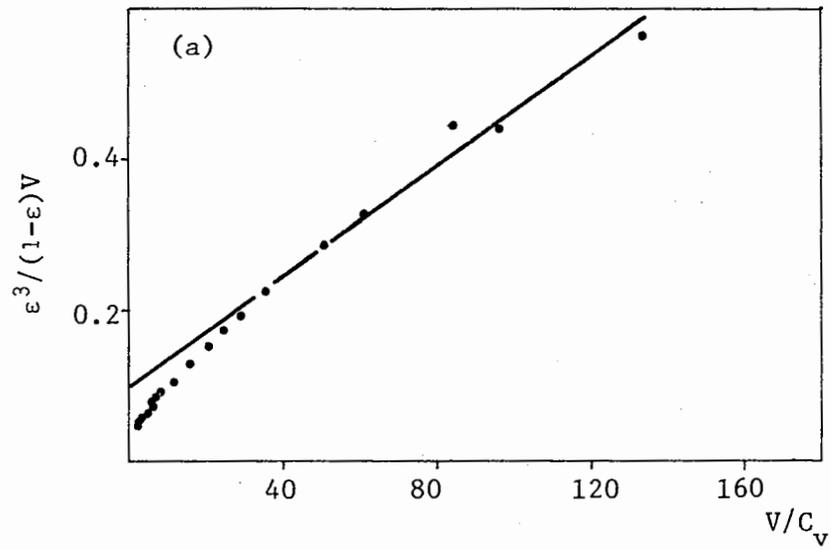


Figure 14. Linearized Mints plot (a) and Kynch curve (b).

(Sacony beads, $d = 0.328$ cm)

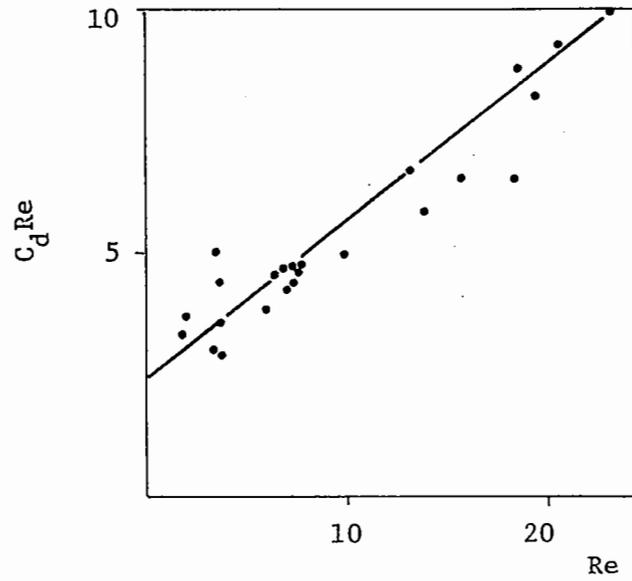


Figure 15. Relationship between resistance coefficient and Reynolds number.

(Granular active carbon, Filtrasorb 400, $d = 0.09$ cm)

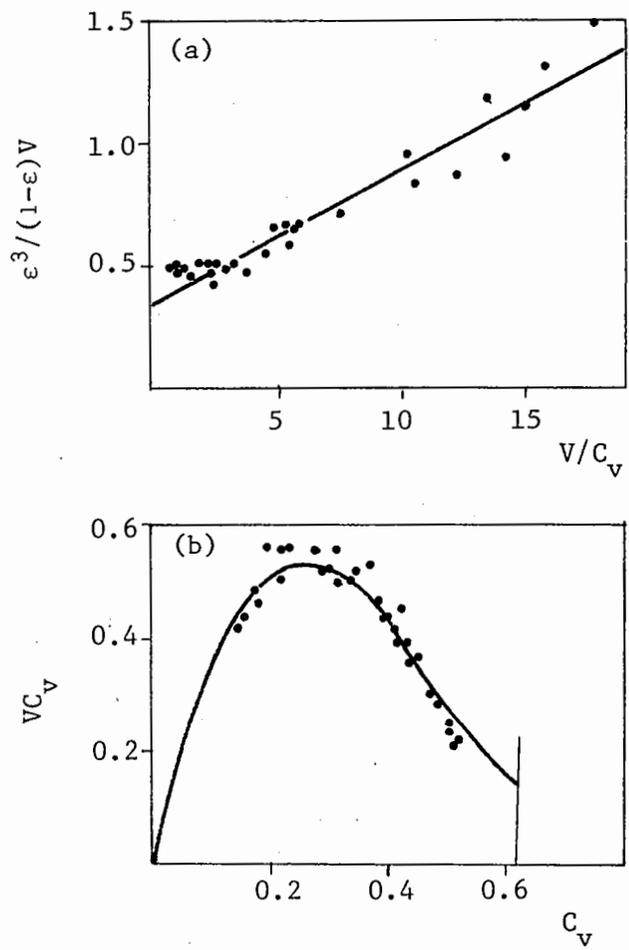


Figure 16. Mints linearized plot (a) and Kynch curve (b).

(Granular active carbon,
Filtrisorb 400, d - 0.09 cm)

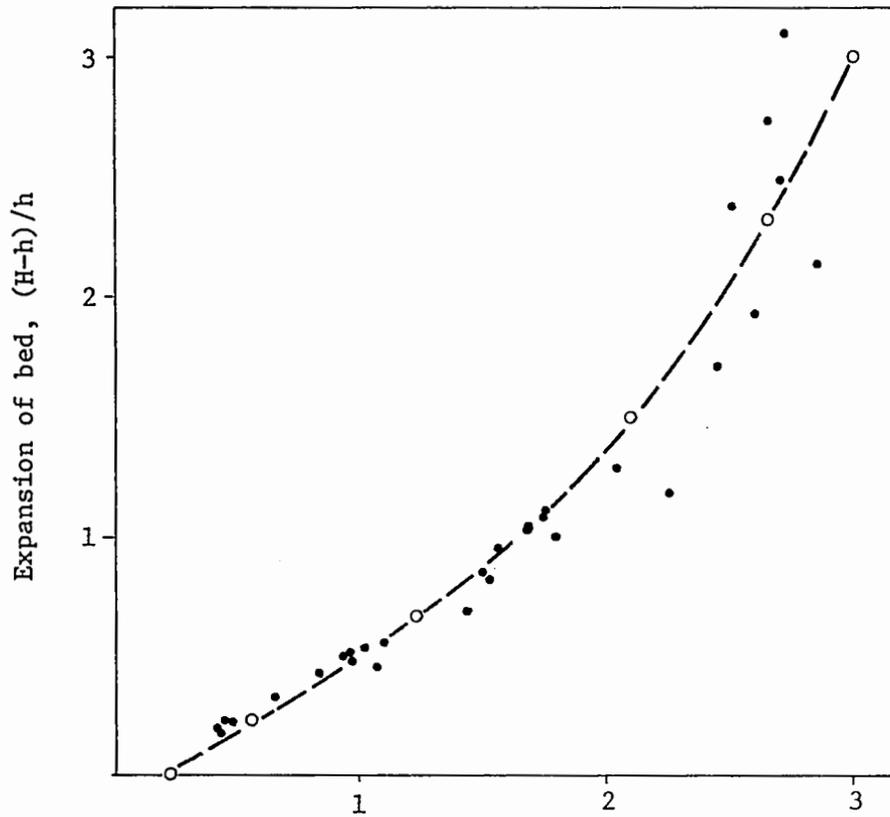


Figure 17. Expansion of the granular active carbon (Filtrisorb 400)

- - experimetnal data
- - predicted by Mints model

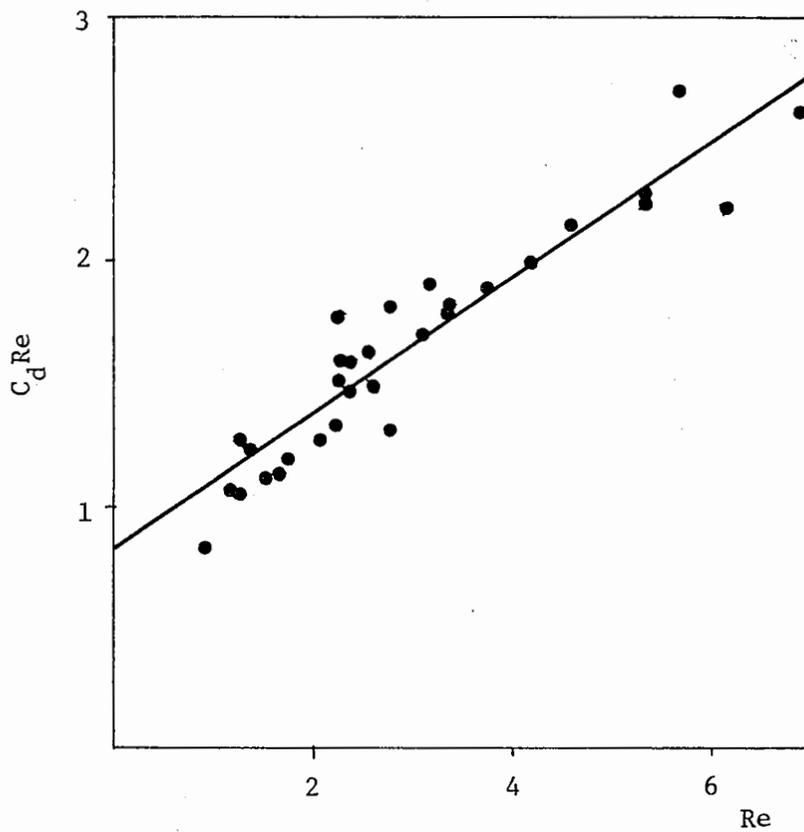


Figure 18. Relationship between resistance coefficient and Reynolds number.

(Black walnut shells, $d = 0.036$ cm)

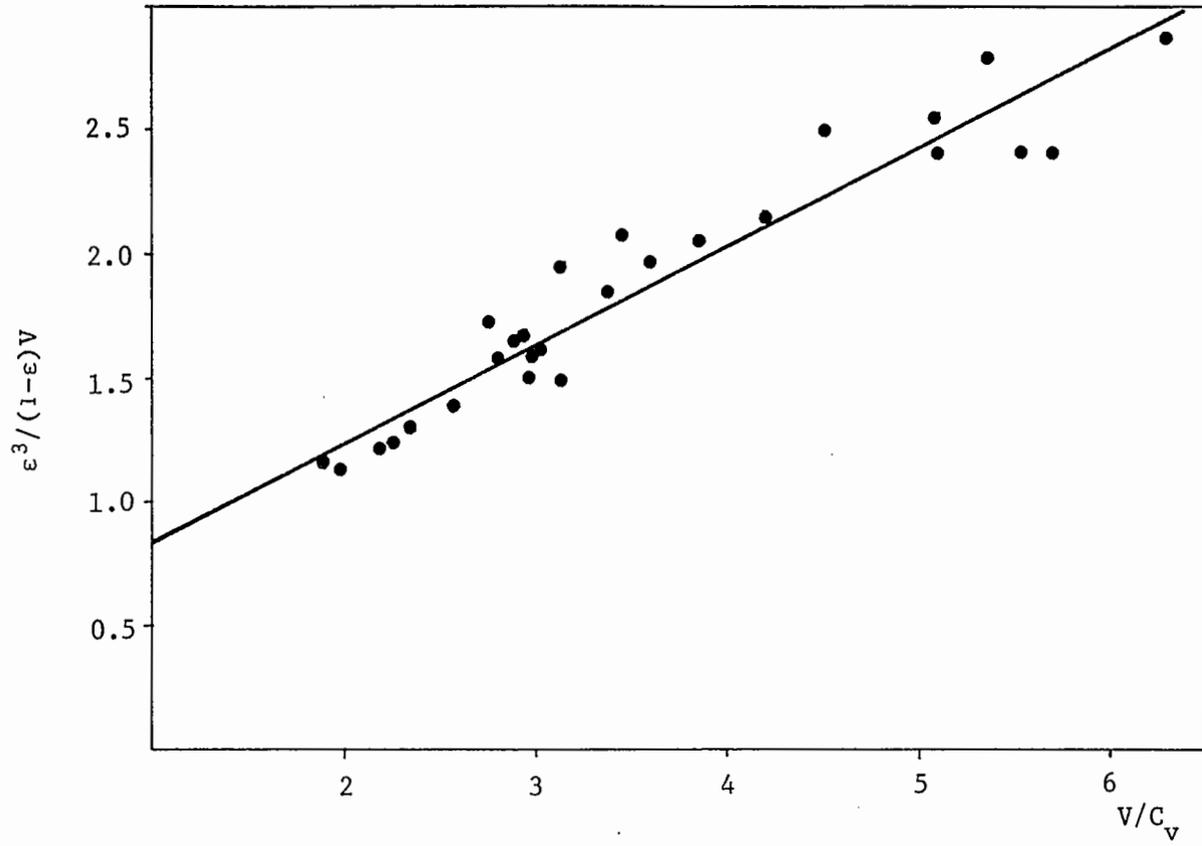


Figure 19. Mints linearized plot
(Black walnut shells, $d = 0.036$ cm)

Table 1. Properties of particles and parameters of fluidization

Parameters	Material									
	Lead Shot	Sand		Glass Beads				Saony Beads	Active Carbon	Black Walnut Shell
d, cm	0.13	0.10	0.04	0.52	0.52	0.52	0.05	0.33	0.09	0.04
$\rho_2, \text{g/cm}^3$	10.8	2.64	2.64	2.35	2.35	2.35	2.49	1.60	1.37	1.24
$\theta, \text{cm/s}$	49.9	12.2	5.5	41.7	41.7	41.7	6.8	19.5	4.48	2.46
Re_p	659	126	21.13	2240	2240	2240	36.2	660	35	9.6
C_p	0.26	0.56	1.01	0.21	0.21	0.21	0.82	0.27	0.85	0.73
ϵ_o	0.38	0.40	0.40	0.39	0.39	0.39	0.38	0.37	0.38	0.39
$V_{mf}, \text{cm/s}$	4.96	0.88	0.132	4.7	4.7	4.7	0.23	1.97	1.15	—

Table 2. Computed parameters of particles and model parameters

Parameters	Material									
	Lead Shot	Sand		Glass Beads				Saony Beads	Active Carbon	Black Walnut Shells
d, cm	0.13	0.10	0.04	0.52	0.52	0.52	0.05	0.33	0.09	0.04
Θ^* , cm/s	50.10	12.4	4.99	43.8	40.6	36.0	4.67	16.7	4.38	2.28
m^{1*}	0.74	1.04	1.19	1.22	0.84	0.54	0.74	0.82	0.78	1.00
m^*	14.0	7.8	2.7	59.8	41.4	26.8	2.33	15.3	2.60	0.75
m^1	0.74	1.02	1.32	1.16	0.87	0.63	1.12	0.96	0.80	1.09
m	13.3	7.66	2.99	57.1	42.7	31.2	3.15	17.9	2.52	0.82
C_o	0.08	0.17	0.40	0.06	0.07	0.09	0.57	0.12	0.28	0.27
Re	$\frac{23}{500}$	$\frac{3.1}{306}$	$\frac{2.9}{56}$	$\frac{79}{10^4}$	$\frac{84}{5150}$	$\frac{74}{2660}$	$\frac{1.38}{48}$	$\frac{167}{760}$	$\frac{1}{23}$	$\frac{1.2}{6.9}$
v	$\frac{5.73}{34.4}$	$\frac{1.04}{9.24}$	$\frac{1.12}{4.84}$	$\frac{4.88}{41.4}$	$\frac{5.18}{36.3}$	$\frac{4.78}{30.1}$	$\frac{0.76}{3.93}$	$\frac{2.11}{13.2}$	$\frac{0.42}{2.85}$	$\frac{0.64}{1.53}$
ϵ	$\frac{0.45}{0.85}$	$\frac{0.43}{0.94}$	$\frac{0.62}{0.95}$	$\frac{0.46}{0.97}$	$\frac{0.45}{0.94}$	$\frac{0.42}{0.88}$	$\frac{0.52}{0.93}$	$\frac{0.41}{0.90}$	$\frac{0.48}{0.85}$	$\frac{0.64}{0.85}$
r	1.00	0.97	0.99	0.98	0.98	0.98	0.99	0.99	0.92	0.95
h_o , cm	10.5	15	10.6	6.2	12	19.5	19.5	17.1	$\frac{3.9}{15.3}$	45

used to determine the settling velocity, θ^* and the coefficient, m^* , where values marked with an asterisk were determined from the linear regression of Equation (12).

The adequacy of the fit of the fluidization data to the linearized form of Mints' model was evaluated by the value of the correlation coefficient. The validity of the model can be judged by the coincidence in the values of θ and θ^* , and m and m^* .

A third set of computations was performed to determine the position of Mints' model on Kynch's curve.

The computed parameters and other supportive data are given in Table 2, and computed lines and curves are shown on the graphs in Figures 2 to 19. In general, the data in Table 2 show good coincidence, with the exceptions of sand ($d=0.04$ cm), glass beads ($d=0.52$ cm, third set), glass beads ($d=0.05$ cm), and sacony beads ($d=0.33$ cm). For glass beads ($d=0.52$ cm), it can be seen in Table 2 that the unexpanded bed height, h_0 , is relatively high compared to the other two sets of data for the same material. This large height relative to the length of the column, 5 feet, permitted only small expansions, $E = (H-h_0/h_0)$, compared to the other two cases. Therefore, most of the data points are located near the regime of the channeled bed. Another factor may possibly be that the particle diameter is fairly large compared to the diameter of the column. In their article, Wilhelm, R. H. and Kwauk, M. (24) indicate that wall effects are neglected. In this case the ratio d/D , where $D = 3$ inch (7.62 cm) = diameter of the column, is approximately $1/15$. Perhaps wall effects are significant at relatively low expansions for larger particles with deeper beds. The same comments apply to sacony beads, $d = 0.33$ cm. Additionally, a random scatter of points for high values of Reynolds numbers, Re , can be seen in Figure 13. This random scatter is the result of the loss of stability of the bed at high expansions, which is characteristic of particles with low densities relative to water, such as sacony beads. For other sets of data which give low coincidences of settling velocity, it can be seen in Table 2 that the correlation coefficients for these materials are very high. Also the corresponding graphs show nearly a perfect linearization of the data, for Reynolds numbers, Re , greater than 1. The authors can only suggest that perhaps an error was made in the experimental determination of the diameter and/or the settling velocities of these relatively small particles. Wilhelm and Kwauk reported in their article that the settling velocity of the sand and small glass beads was obtained in a

500 cc cylinder, instead of the 3 inch diameter column. This is a deviation in experimental procedures, which supports the authors' opinion that the physical properties of small particles are difficult to determine by the counting and weighing method (for determination of particle diameter) and dropping particles to determine the settling velocity of the particles.

STRUCTURE OF THE LIQUID-SOLID FLUIDIZED BED

Some older studies and more recent studies using sophisticated measurements have shown that beds fluidized by liquids, which visually appear to be uniform, actually contain porosity fluctuations. Kurguev, E. F. (12) detected periodic oscillations of velocities and significant fluctuations in the concentration of suspension in suspended sludge blanket clarifiers. Lawson, A. and Hasset, N. J. (14), using cinematographic techniques, showed the propagation of parvoids, consecutive denser and diluted layers in the bed. Vanacek, V. and Hummel, R. L. (23) detected periodic and non-periodic variations in instantaneous velocities in fluidized beds. These are only a few of the articles, that describe non-uniformities in the structure of fluidized beds. The authors have observed that the level of the fluidized bed oscillates with an amplitude of several millimeters. These oscillations are usually accompanied by weak surface waves. In general, the presence of an extremely thinned layer at the base of the bed and horizontal waves propagating along the height of the bed are indicated.

The authors suggest that an ideal model can be helpful for studying the problem of hydrodynamic instability in fluidized beds. The deviation of measured parameters from the ideal model could be used for evaluation of the degree of instability of the system. Based on the authors' observations and the articles previously referenced, the following structure of the fluidized bed can be suggested.

Description of Fluidization

The changes in physical characteristics of a bed of granular material, when subjected to small stepwise increases in the upward flow rate are discussed. Initially the liquid flows through the pores between the solid particles. The

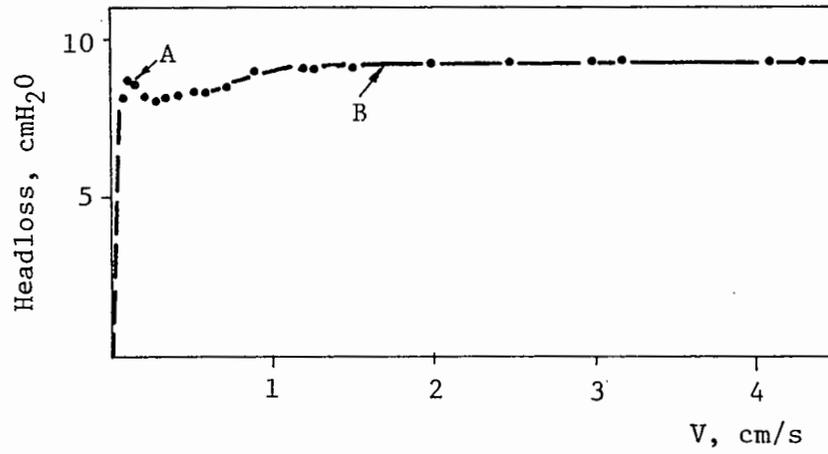


Figure 20. Headloss across the fluidized bed as a function of velocity of flow.
 (Sand, $d = 0.37$ cm)

headloss is proportional to the superficial velocity of flow with the bed retaining its original compacted configuration. This corresponds to the dashed line portion between the origin and point A on the graph in Figure 20 and the vertical line in Figure 7b. A point is reached when the increase in flow causes a rearrangement of the particles within the skeleton as the particles rearrange into a loosened structure which causes the least resistance to flow. For this range of flows the bed will begin to expand and channels of liquid will develop, corresponding to the data points between A and B in Figure 20, where a reduction in headloss occurs, and in Figure 7b where a non-linear increase in the solids flux, due to expansion of the bed with increased velocity. In this range of points the bed is expanded, but not fluidized. A sharp discontinuity occurs, for increased flows. This discontinuity is apparent on the graphs in Figures 6, 7 and 20. Beyond this point the data coincides extremely well for all of the curves determined by Mints' model. For increased flows the bed continues to expand, in accordance with Mints' model, in order to keep the hydraulic resistance at a minimum, equal to the submerged weight of the bed.

It is apparent from the graphs that a critical condition occurs for Reynolds numbers approximately equal to 1. This condition is noticeable for small materials and/or materials with low densities. There are no data points for Reynolds numbers less than 1 in Figure 15 for granular activated carbon and a single point (at $Re=0.93$) for black walnut shells in Figure 18. These data were obtained by the authors who were unaware of the critical condition at the time the data was collected. The absence of data for this regime can be explained by the fact that measurements were only taken when the bed visually appeared to be fluidized.

Based on the consistent data discussed above, the authors suggest that a minimum Reynolds number exists, below which the condition of fluidization for liquid-solid systems does not exist. This represents a lower limit to the validity of Mints' model.

Ideal Fluidization

The following description of the structure and dynamic nature of the ideal state of fluidization is suggested. Ideal fluidization occurs through a lifting of the bed from the support and subsequent precipitation of the particles from the lower layer of the bed into the thinned liquid layer

beneath the bed. Precipitation of particles accelerates the flow of water, due to the effect of displacement. As a result, the settling rate of the "second wave" particles departing from the bottom of the uplifted dense layer is lessened. The second wave particles form a more concentrated band than the first one. Liquid filtering through the second wave lifts the non-fluidized bed further, resulting in the formation of a layer of lower density over the previously formed denser layer. The process continues until the entire bed is transformed into a sequence of low and high density layers.

Particles from the lower boundaries of dense layers precipitate across the layers of low density and onto the top boundaries of dense layers. This means that the dense layers gain material from the top and lose materials from the bottom. This mechanism results in the upward propagation of denser layers. The layer at the base of the bed always remains diluted. However, its thickness oscillates in accord with the propagation of waves across the bed. This mechanism of fluidization assumes a discontinuity in the concentration of particles along the height of the bed. Such discontinuities are in accord with Kynch's theory of hindered sedimentation.

Deviations from Ideal Fluidization

Gelperin, N. I., Ainshtein, V. G., and Kwasha, V. B. (5,6) have shown that the fluidized layer exhibits properties of a liquid. It follows from the previous discussion that the fluidized bed can be considered as a system of liquid layers of greater density underlain by liquid layers of lower density. Such a system is discussed by Monin, A. S., and Yaglom, A. M. (18) as a classic example of an unstable system. Under the action of small disturbances these layers tend to flip over. In a running process this causes a redistribution of pressure at the bottom of the reactor and self-sustained boiling of the bed. The redistribution of pressure is more significant in reactors of larger dimensions, and consequently, in these reactors boiling is more likely to occur.

The idealized model does not consider the cohesion and interlocking of particles in unexpanded beds. If these factors are significant, the liquid may form vertical channels through the bed. This situation is typical for smaller and more flocculent particles at low velocities of

flow.

Boiling and channeling occur due to internal properties of fluidized systems. Bed stability depends on water distribution devices only as much as it concerns the induction or reduction of small disturbances, which can cause the loss of stability. Only a means located within the bed can exert control over the finer parameters of the system, thereby increasing the stability of the bed. One of such means is discussed in the following section.

PRELIMINARY STUDY OF THE EFFECTS ASSOCIATED WITH FIXED PACKING

Considering the fact that instabilities of both kinds, channeling and boiling, are associated with nonuniformities of velocity and concentration of particles; a device that produces mixing in the bed will equalize the distribution of particles and velocities and increase the stability of the bed. Devices such as this should not induce major flow streams in the bed. The authors suggest a special type of fixed packing, which produces localized mixing in the bed. One possible configuration of the packing is presented in Figure 21. The packing is comprised of one or several horizontal tridimensional girders, consisting of slanted plates which form alternating contracting and expanding cells.

A preliminary experimental study of the effects associated with fixed packing was performed using the columns, which were described previously. The small column was fitted with a single slanted plate, simulating the simplest form of the fixed packing. The dimensions of the plate were 8 cm (length) x 0.65 cm (thickness) x 1.27 cm (width). The plate could be rotated 360° and fixed at any desired angle. Thirty plates with dimensions 5x0.65x1.27 cm³ were placed in the large column. They were placed in 5 rows of six plates at an angle of 15° with the vertical, alternating the angle from left to right, thereby creating expanding and contracting cells. A particular plate in a row was directed in opposition to the plate directly above and below it. The plates were evenly spaced across the width of the column. The rows were located at heights from 20 cm to 110 cm above the base of the column. This configuration permitted examination of conditions of fluidization either totally within the zone of plates or expanded to some height beyond the plates.

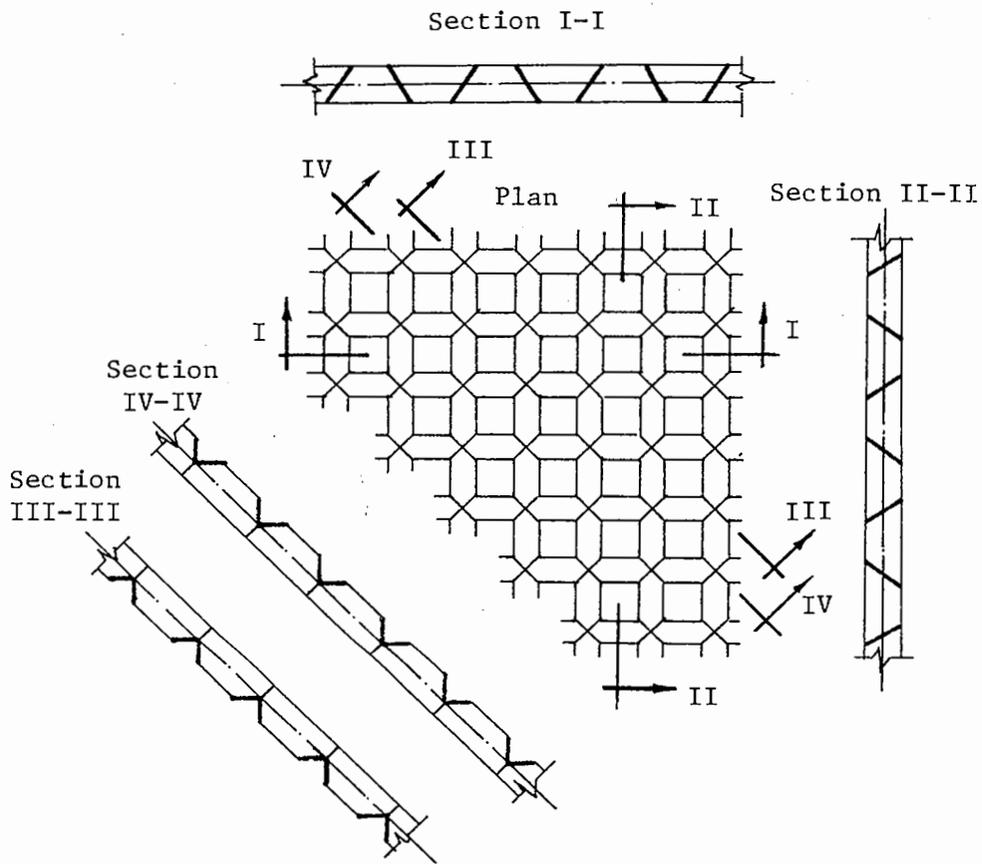


Figure 21. Suggested design of fixed packing.

Flow Patterns With the Fixed Packing

Rotation of liquid and solid material around the plates was observed. The rotational flow is directed upward in contracting cells, across the upper tip of plates, downward along the plates, and toward the contracting cells at the lower tip of the plates. A liquid boundary layer is formed on the bottom side of the plates. A layer of solid particles sweeps along the top side of the plates. This layer is formed by particles which precipitate in the expanding cell. The thickness of this layer and the velocity of this layer are not constant in time. The suspension concentrates in the expanding cell and is periodically swept to the contracting zone. This pulse induces a shock regime of flow in the contracting cell. The intense rotation of liquid and solid material around the plates and the shock regime generated by this rotation results in mixing of the solids within the girder and the dissipation of energy. It is assumed that mixing and the dissipation of energy enhance the stability of the bed, through insuring more uniform patterns of flow and more uniform distributions of concentration.

Discussion of Preliminary Experimental Data

The small column was used for the observations of flow patterns around a slanted plate and for measurement of the effect of a single plate on the expansion behavior of the fluidized bed. Figure 22 illustrates the effect of varying the angle of attack of the plate on bed expansion. For a comparatively large range of velocity and for all angles of attack, the expansion of the bed with the plate was lower, compared to the bed without the plate at the same velocity of flow. At angles of attack greater than 25° to 30° (with vertical) a stationary layer was formed on the topside of the plate. This layer was held in place by local hydrodynamic pressure at an angle much greater than the angle of repose of sand. This indicates that hydrodynamic pressure differentials at the plate are considerable. However, it is important to stress that the formation of deposits in fluidized bed reactors is not desirable. Therefore, limitations on the angle of attack must be established.

Measurements obtained from fluidization experiments in the large column are presented in Figures 23 to 25. Relationships between bed expansion and fluidization velocity for black walnut shells in the large column with and without

fixed packing are illustrated in Figure 23. The experimental data for tests without fixed packing are modeled by Mints' equation. Deviations of the data points from the ideal model are greatest at high velocities of flow, due to bed boiling. Comparison of expansion for tests with and without fixed packing shows that for the same velocity, the expansion is greater for the column with fixed packing when the fluidized bed is totally within the packed zone. However, the overall expansion for tests with fixed packing indicates a downward trend when the expanded height of the fluidized bed is above the level of the fixed packing. At a velocity of flow; $V = 1.18$ cm/s, the expansions of the bed are about equal. At greater velocities, the expansion of the bed without fixed packing is greater, than that of the bed with fixed packing for the same velocity.

Both phenomena, the increased expansion within the fixed packing and the decreased overall expansion have practical applications, for controlling the expansion behavior of the bed. It can be seen for velocities greater than approximately 1 cm/s, the scatter of points for the column without packing is greater than that for the column with packing. It was observed that the bed with fixed packing was less susceptible to boiling at high velocities. In the low range of velocities, fluidization could be achieved at a lower velocity in the column with the fixed packing. Therefore, the amount of water required for fluidization is reduced.

Figure 24 presents Kynch's curve computed on the basis of Mints' model using experimental data points obtained in the column without fixed packing. It should be noted that relatively stable fluidization for the black walnut shells was limited to the range of volumetric concentrations, C_v , from 0.15 to 0.4. In the column with fixed packing the solids flux obtainable at low concentrations is higher than that obtainable from the "standard" Kynch curve. At higher concentrations, stable fluidization occurs at lower velocities, than in the column without fixed packing.

Figure 25 presents data on the expansion rates of the bed for three corresponding velocities of flow. The graphically determined time constants for this transient process are presented in Table 3.

It follows from Table 3 that transient processes occur faster in columns with fixed packing. The packing results in more energy per unit time spent for expansion. Therefore, the magnitude of the disturbance applied to the column with packing which will induce instability must be greater than

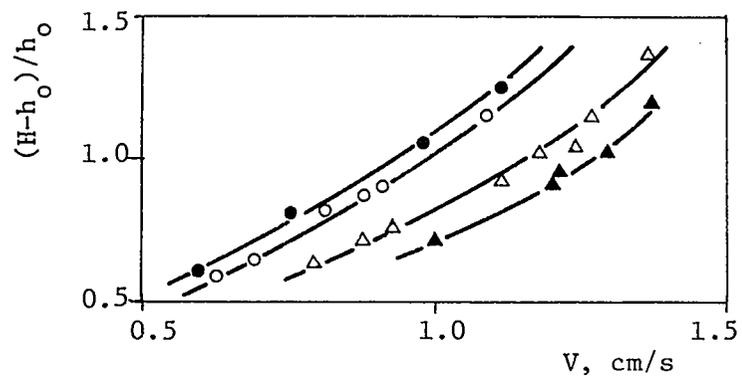


Figure 22. Expansion of the bed with a single plate (small column)

(Sand, $d = 0.032$ cm, small model)

● without plates; ○ 15° angle of attack; △ 30° ; ▲ 45° .

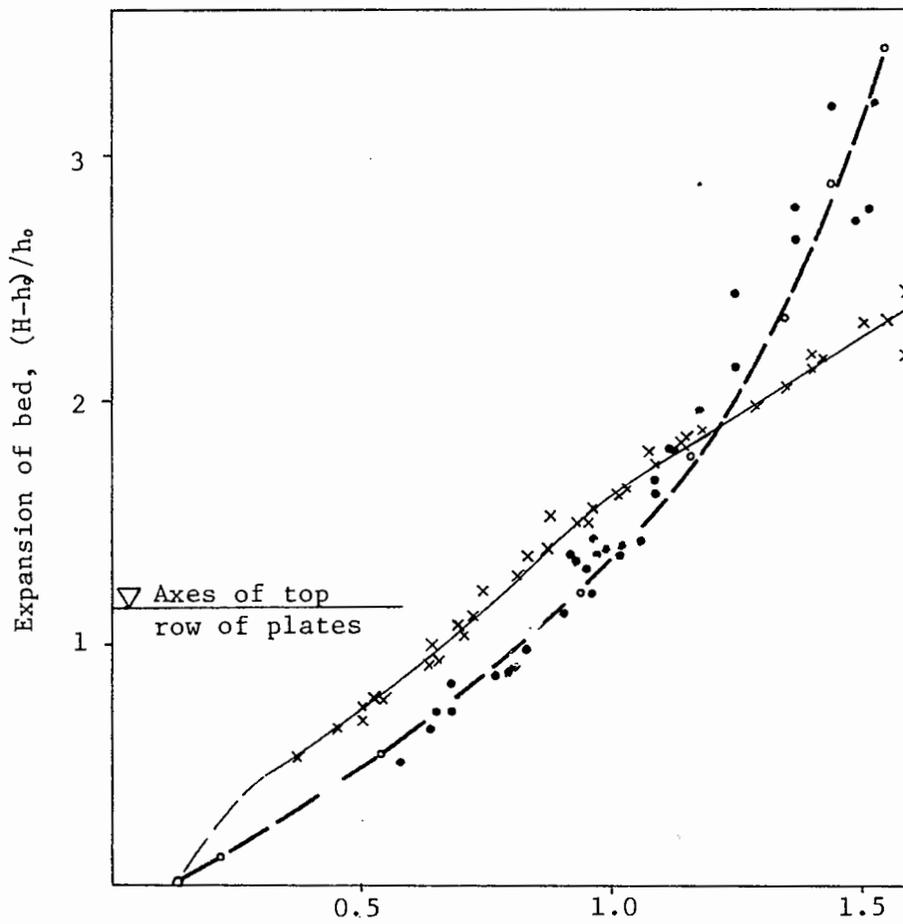


Figure 23. Expansion of the bed without plates (\cdot), and with plates (\times).

\circ - expansion of bed without plates predicted by Mints model

(Black walnut shells, $d = 0.036$ cm, large model)

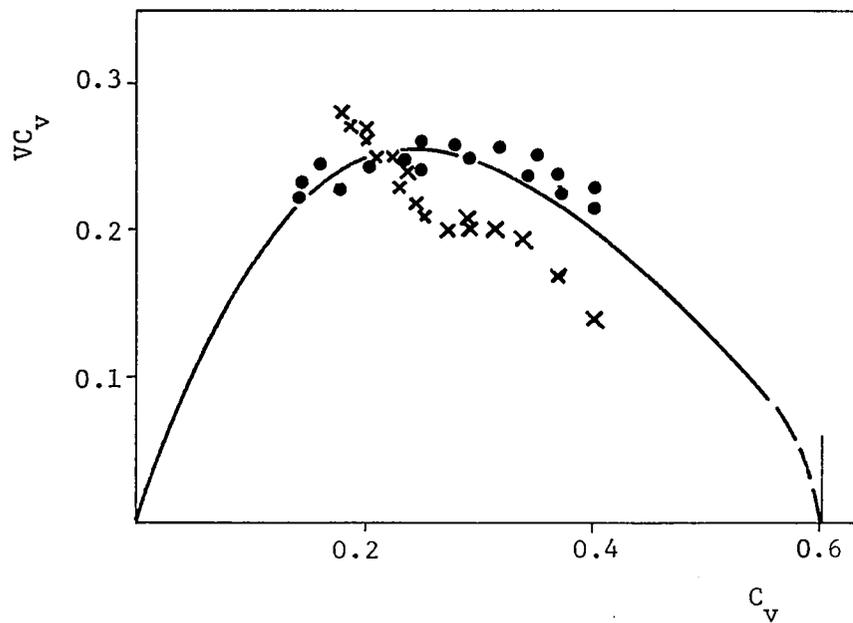


Figure 24. Kinch curves for columns with (x) and without (•) plates.

(Black walnut shells, $d = 0.036$ cm, large model. Solid line shows Kynch's curve computed by the use of Mints model.)

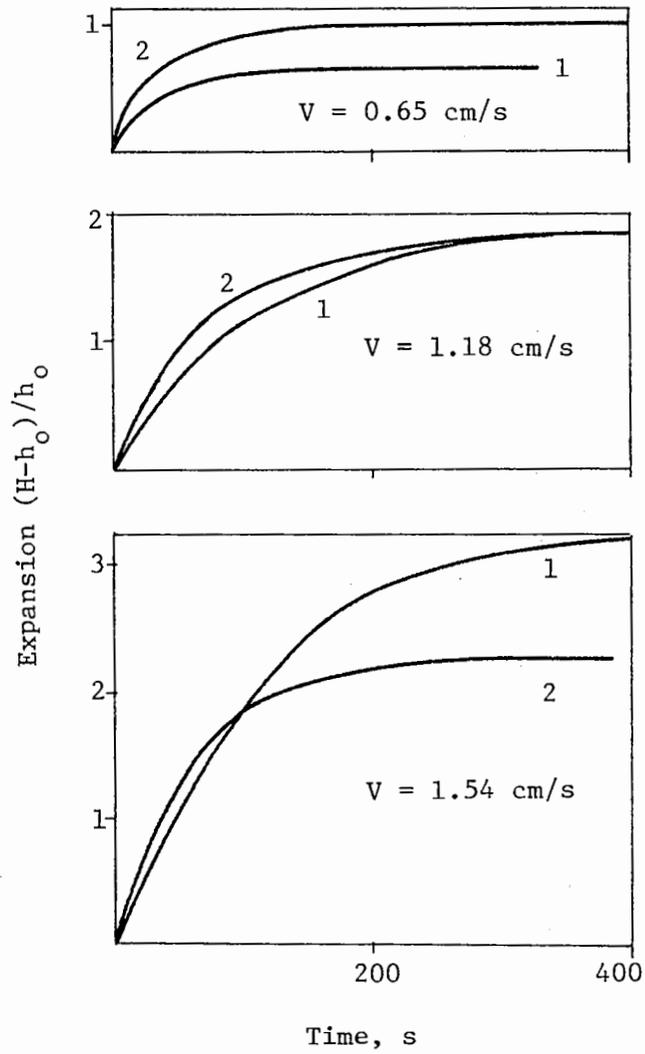


Figure 25. Expansion rate curves

(Black walnut shells, large model)

1. without packing,
2. with packing.

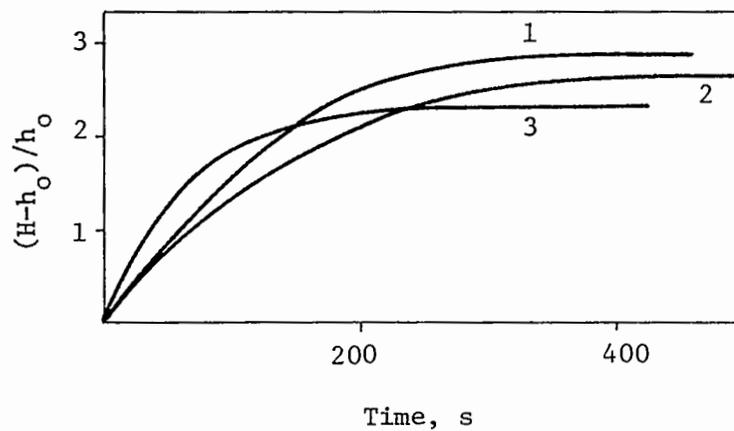


Figure 26. Rate of expansion of the bed (Black walnut shells, large model)

1. without packing, $V = 1.37$ cm/s (boiling regime, four inlet holes plugged),
2. without packing, $V = 1.37$ cm/s (stable regimes, all holes operable),
3. with packing, $V = 1.42$ cm/s (stabilized regime, four inlet holes plugged).

that applied to the column without packing.

Table 3. Time constants for expansion of the bed in columns with and without fixed packing

Column Type	Velocity of flow, cm/s		
	0.65	1.18	1.54
1. Without packing	53.0	60.2	45.5
2. With packing	24.1	47.9	30.5

Figure 26 compares the transient expansion process at comparable velocities of flow for:

- (1) induced non-uniform flow distribution without packing, resulting in a boiling bed,
- (2) uniform flow distribution without packing, resulting in stable fluidization at the end of the transient process,
- (3) induced non-uniform flow distribution with fixed packing, resulting in stable fluidization at the end of the transient process.

The final expansion for the boiling bed (1) is greater than that for the stable bed (2). The column with fixed packing (3) produced a lower expansion with induced non-uniformity of flow than case (2). The fixed packing stabilizes the bed and retains the expansion at same level as occurred in the case of uniform water distribution (see Figure 23).

The data on the effects associated with fixed packing show that channeling and boiling can be reduced. When this packing is developed, it will have practical applications to fluidized bed biological treatment processes and other treatment processes, such as: filtration, sludge separation and thickening, mixing, etc.

CONCLUSIONS

Mints' model adequately describes ideal static

(nonvariable input parameters) liquid-solid fluidization. This model has a good theoretical basis and describes bed expansion relationships for a wide range of experimental data. This model has been verified for materials with low densities relative to the fluidizing liquid. The lower limit for the validity of Mints' model occurs at a Reynolds number, Re , of approximately 1. It is suggested that below this value the state of fluidization does not exist, however, the regime of flow may correspond to an expanded bed. Deviations from the idealized model occur at low and high Reynolds numbers. These deviations correspond to unstable flow regimes, exhibited in the form of channeling and boiling. The loss of stability at high Reynolds numbers is due to the inherent instability of the layered structure of liquid-solid fluidized beds.

A fixed packing consisting of one or more girders creating expanding and contracting cells is proposed to increase the stability of fluidized beds. Preliminary experimental data has been presented, which indicates that this fixed packing appears promising. More experimental work is needed to develop the concept and determine design parameters.

NOTATIONS

C = coefficient,
 C_o = resistance coefficient,
 d^d = diameter of particles,
 E = fractional expansion,
 Ga = Galileo number,
 H = height of fluidized, or expanded, bed,
 h_o = height of unexpanded bed,
 ℓ_o = hydraulic radius of porous space,
 m = coefficient (characteristic of particles),
 m^1 = dimensionless characteristic of particles,
 m_f = subscript for minimal fluidization,
 P^f = hydraulic resistance of the bed,
 Re = Reynolds number,
 u = true velocity of flow in porous space,
 V = velocity of flow in "empty" tank,
 β = ratio of hindered settling velocity, V , to the terminal settling velocity, θ ,
 ϵ_o and ϵ = porosities of unexpanded and expanded (fluidized) beds,
 θ = terminal settling velocity of particles,

ρ_1 and ρ_2 = densities of liquid and solid phases,
 μ = viscosity,
 ω = grain surface per unit volume of bed, $\omega = 6(1-\epsilon)$,
* = is used for computed values.

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both sides of the equation by θC_v , an expression describing Kynch's curve, flux versus concentration, is obtained. This equation contains the terminal velocity of a particle, θ , and an empirical parameter, m^1 . In order to facilitate the determination of these parameters, equation (10) can be linearized by substituting the relationship $C_d = C_p/\pi$ into (10), multiplying it by πV , and dividing by $C_p(1-\epsilon)\theta^2$. This results in the following expression:

$$\frac{\epsilon^3}{(1-\epsilon)V} = \left[\frac{1}{\theta^2}\right] \frac{V}{(1-\epsilon)} + \left[\frac{2m^1}{\theta}\right] \quad (13)$$

This expression relates the measurable parameters, V and ϵ . Plotting the experimental values in $y = \epsilon^3/[(1-\epsilon)V]$ versus $x = V/(1-\epsilon)$ produces a straight line with a slope, $1/\theta^2$, and the segment cut on the ordinate $2m^1/\theta$. The plot, in accordance with equation (13), is especially convenient when particle characteristics, such as diameter, shape factor, etc. are unknown.

VERIFICATION OF THE IDEAL MODEL

It was stated in the previous section that Mints verified the model for large and relatively dense materials. However for fluidized bed biological reactors it is desirable to use relatively small and light particles, because of the advantages of the greater specific surface area and relative ease of fluidization afforded by these particles. For this reason experiments were performed with "small" sand, granular activated carbon, and crushed black walnut shells to verify Mints' model for these materials. Additional data was obtained from the literature, Wilhelm, R. H. and Kwauk, M. (24). The experimentally determined physical properties of the particles is presented in Table 1. The specifics of the experimental apparatus and procedures are described in the following subsection.

Experimental Installations and Procedures

Two rectangular plexiglass columns with internal dimensions 60 x 10 x 1.27 cm³ were used by the authors. The columns had a thin third dimension to enhance visual observations of the fluidized bed through backlighting. A schematic of the small experimental column is presented in Figure 1.

RETENTION AND DISTRIBUTION OF BIOLOGICAL SOLIDS IN FIXED-BED ANAEROBIC FILTERS

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INTRODUCTION

A number of research investigations conducted during the past ten to fifteen years have shown that the fixed-bed anaerobic filter process is ideally suited for treatment of low-strength soluble wastes. In this process, wastes are passed upward through a fixed or stationary bed of porous media (Figure 1). Biological solids become attached to the surfaces of the media or are trapped in the void spaces in high concentrations so that the long solids retention times (SRT) necessary for satisfactory anaerobic treatment of organic wastes at nominal temperatures are attained. Settling and recycle of the effluent solids are not required to maintain a high treatment efficiency, and solids separation is not required as a means of process control. With the low net solids synthesis, nutrient requirements are reduced and solids disposal problems are minimized. Since the system operates fully submerged, the hydraulic head requirement is low.

The anaerobic filter is a "fixed-film" process in which the stabilization of wastes takes place at the surface of a layer of biological solids attached to or held by the filter medium. Since flow through the filter approaches plug-flow,

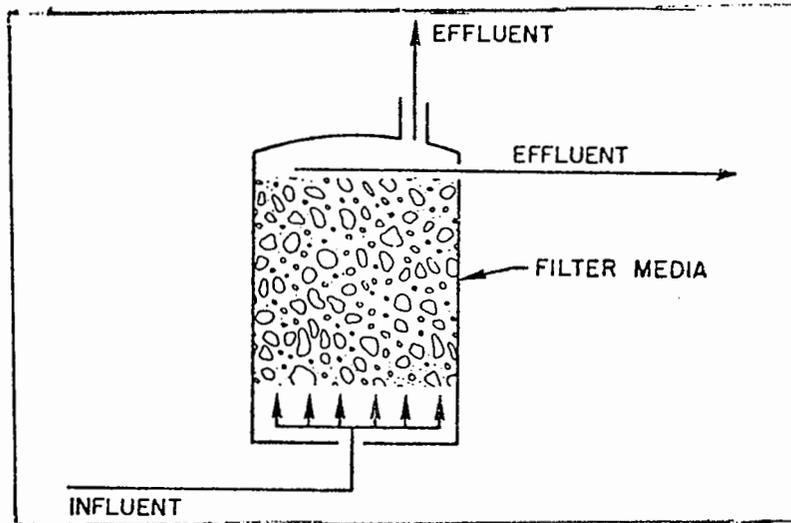


Figure 1. Schematic diagram of the anaerobic filter process

a high degree of biological efficiency is approached. A high rate of removal occurs in the lower levels where both substrate and biological solids are present in high concentrations. As the waste flows through successive layers of the filter, organic material is removed by the active biological solids in that layer.

The mode of operation of the anaerobic filter is somewhat analogous to that of a number of reactors in series with high-rate treatment in the first unit and polishing and solids separation in the following units. The major advantages of such multi-stage anaerobic processes are 1) that an effluent can be produced having much lower concentrations of unutilized substrate than is possible in completely mixed systems, and 2) environmental conditions in each stage may favor the development of a particular group of microorganisms and thereby increase the rate of consumption of a particular component of a waste.

A knowledge of the factors affecting the retention and distribution of biological solids throughout the reactor height is essential to an understanding of the performance capabilities of anaerobic filters and to their proper design and application. The purpose of this paper is to identify some of these basic factors and to show by use of laboratory tests how these factors affect waste treatment performance in plug-flow, fixed-bed anaerobic filters.

BACKGROUND STUDIES

McCarty (1), in 1966 first suggested the use of fixed-film anaerobic processes for industrial waste treatment when reporting the results obtained from a three-liter laboratory-scale filter containing 25 to 37 mm diameter quartzite stone. For 226 days of operation at a 12-hour theoretical hydraulic detention time, the COD removal efficiency averaged 81 percent. Effluent suspended solids were normally below 20 mg/l and no solids wasting was required throughout the period of operation.

A comprehensive laboratory study subsequently was conducted by Young (2) and reported by Young and McCarty (3) to investigate the potential for using anaerobic filters to treat low-strength soluble wastes and to observe their treatment performance when operating over a wide range of loading conditions. During this study, eight laboratory-scale filters, each 0.14 m in diameter, 1.83 m in height, and having a volume of 28.5 liters, were filled with 25 to 37 mm diameter smooth quartzite stone. The porosity of this medium was 0.42. Synthetic waste consisting of an equal-COD mixture of protein and carbohydrates was fed to these filters at concentrations ranging from 1,500 to 3,000 mg COD/l. Organic loadings ranged from 1 to 8 kg COD/day per cubic meter of void volume when operating at 25°C.

Numerous other studies followed with objectives ranging from investigation of fundamental operating characteristics to development of criteria for designing full-scale facilities (4-6).

One question arising from these studies concerned the effect of media type, size and shape on filter performance. Consequently, the authors in 1980 began a series of investigations to identify media-related factors affecting the treatment performance of fixed-bed anaerobic filters (7). These studies involved the use of a number of different types and sizes of media in large laboratory-scale reactors 0.5 m in diameter and 2 m in height. These large reactors permitted the use of commercially available media, as shown schematically in Figure 2, with minimum wall effects. Two reactors were filled with modular blocks constructed from cross-stacked corrugated plastic sheets (Models 19060 and 27060, The Munters Corp., Fort Myers, Florida). A third reactor contained Pall rings 90 mm high by 90 mm diameter (Actifil, The Norton Company, Akron, Ohio). Size and other media characteristics are listed in Table I.

Table I. Media Type, Size and Shape Characteristics

Media Type	Size, Description	Porosity	Unit Surface Area, m^2/m^3	Average Pore Diameter, mm
Rock	25 mm to 37 mm quartzite stone	0.42	20 to 30	12
Pall Rings	90 mm x 90 mm (Norton Actifil, No. 90E)	0.95	102	20
Large Modules	Corrugated sheets; openings = 50 mm x 80 mm (Munters No. 27060)	0.95	98	46
Small Modules	Corrugated sheets; openings = 20 mm x 40 mm (Munters No. 19060)	0.95	138	32

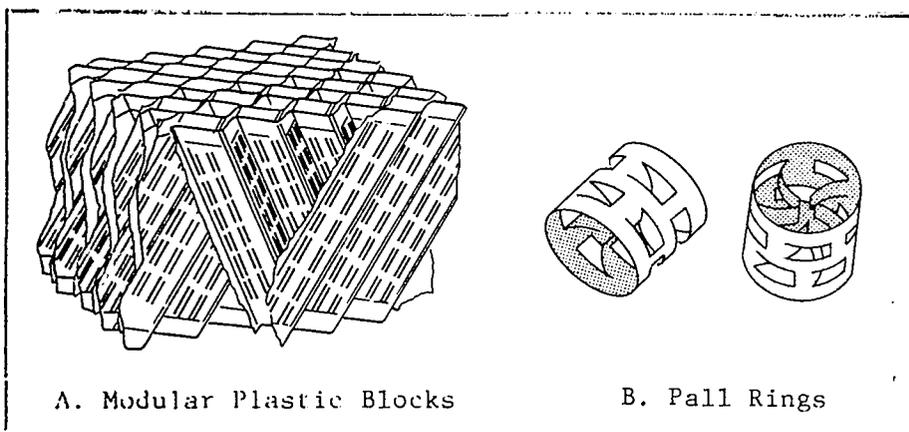


Figure 2. Schematic diagrams of the two types of plastic media used in the laboratory-scale anaerobic filters.

The waste fed to these units was a synthetic alcohol stillage containing a mixture of sucrose (29.2% COD by weight), volatile acids (5.5% COD by weight), and alcohols (65.2% COD by weight). This waste was fed to the units at organic loadings ranging from 0.5 to 16 kg COD/day per cubic meter of reactor void volume.

TREATMENT PERFORMANCE

The test units for both the investigations described above were operated after seeding for sufficient lengths of time for their treatment performance to reach "steady-state" as defined by overall constant COD removal efficiency and constant gas production rate. Samples were collected at various reactor heights throughout the period of operation and were analyzed for residual COD, suspended solids and volatile organic acids to provide an indication of the treatment performance as related to height.

COD Removal Performance

As shown in Figure 3, when operating at an organic loading of 2 kg COD/day per cubic meter of void volume and an influent COD of 1500 and 3000 mg/l, the rate of COD removal was quite high in the first 0.3 m of filter height. There were only slight differences in the shape of the COD profiles when operating with the same influent conditions. While the rock medium seems to have provided better overall COD removal and more gradual change in COD with height than did the other media, it is difficult to interpret the significance of this difference because the reactor containing the rock was operated at 25°C and with an influent waste that was different from that fed to the other reactors.

The rapid decrease in COD does, however, suggest that factors affecting the concentration of biological solids and removal of wastes are most important in the first 0.5 m of height.

Biological Solids Distribution

After a period of operation over a range of successively increasing loadings, at least one reactor containing each of the four types of media--rock, two sizes of plastic modular media and plastic Pall rings--was dismantled. The medium

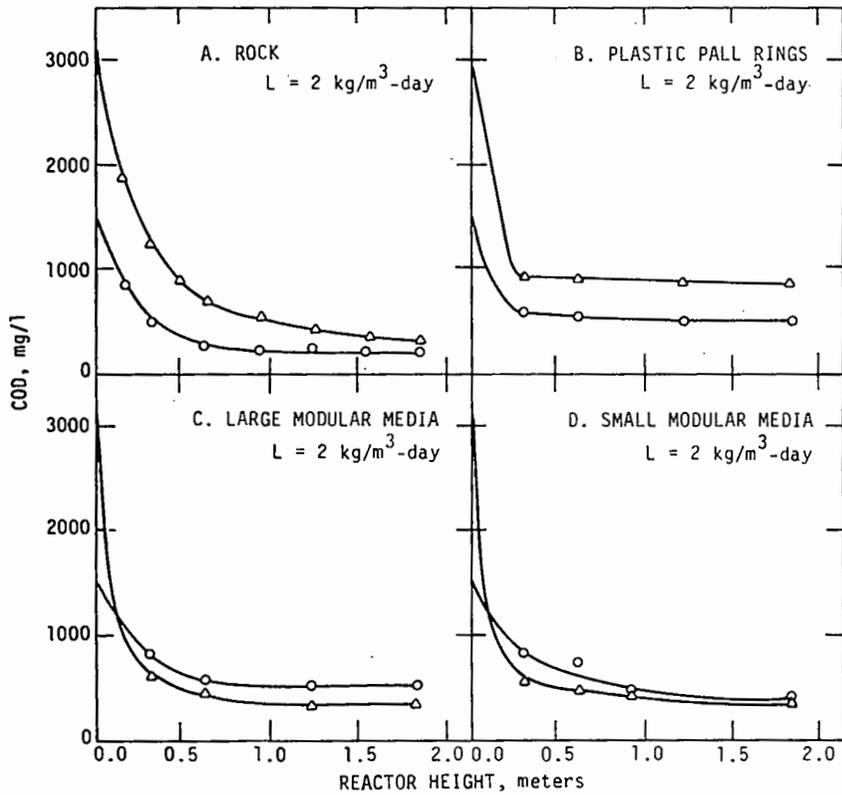


Figure 3. COD profiles throughout anaerobic filters containing four different media and when operating at an organic loading of $2 \text{ kg COD/m}^3\text{-day}$.

from each reactor was removed carefully in sections and the attached solids were removed. Suspended growth was removed by siphoning the liquid from each section as the media modules were removed.

All solids measurements were converted to a mg/l concentration in the void volume of the reactor at the height

section from which the solids were removed.

The association of rapid COD removal with high concentrations of biological solids is verified by comparison of the COD profiles in Figure 3 with the solids distribution profiles in Figure 4. The loading at which each reactor was operated prior to terminating its operation is given beneath

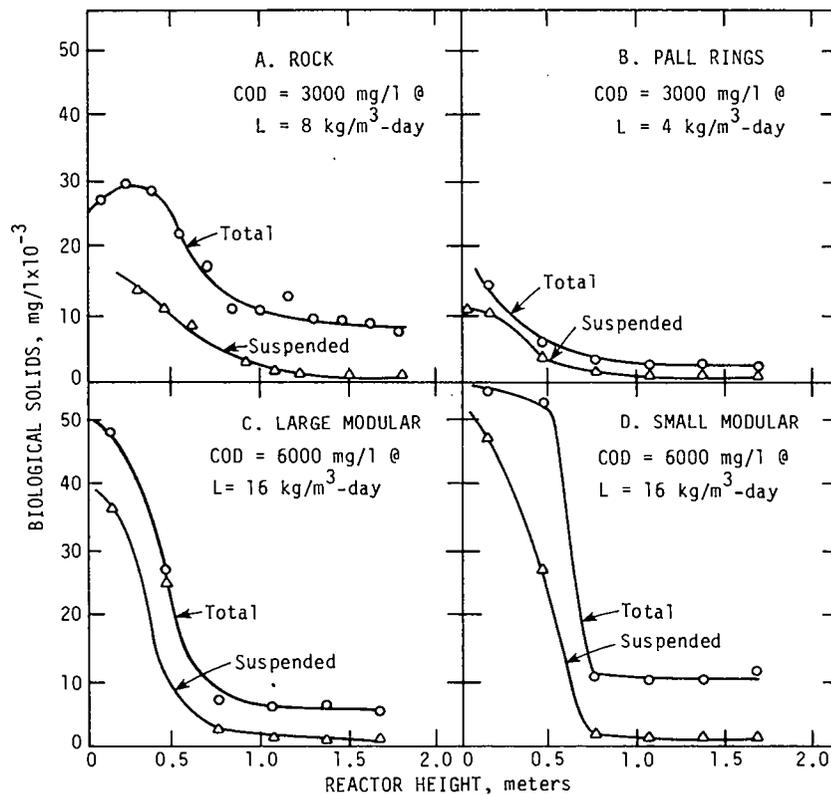


Figure 4. Biological solids distribution in anaerobic filters containing four different media.

the heading on each quadrant of Figure 4. These differences in the schedule and duration of the loadings preclude direct comparison of biological solids concentrations between all reactors. However, the data do show that the relative distribution of attached and suspended growth was similar regardless of the media type, size and shape.

A large part, usually one-half to two-thirds of the total solids mass was held loosely in the interstitial void spaces. The remainder was considered to be attached in that it could not be withdrawn by gravity drainage and would not fall from the media modules when they were lifted gently from the reactors.

Indication of the activity of these solids is provided in Figure 5. Figure 5B shows the rate of COD removal per unit of total biological solids mass for reactors containing the rock and the three plastic media described in Table 1 when operating at loadings identified with the respective COD profiles in Figure 5A. The decrease in COD removal rates at the upper filter heights as compared to that at the lower heights is evidence of the lower effectiveness of the biological solids at the upper reactor heights.

Further insight into the activity of the biological solids in the test reactors was gained by conducting serum bottle tests using the suspended growth removed from the unit containing the small modular medium. Known amounts of suspended solids were placed into sterile (250 ml) serum bottles that had been flushed with nitrogen gas. After a 12-hr period of adjustment, a measured amount of the synthetic stillage was added to each of a number of the serum bottles. Methane production was monitored using a procedure described by Johnson (8). The maximum rate of methane production in the test bottles receiving the waste material, as compared to the maximum rate of production in control units receiving no waste, indicated the activity of the biological solids.

As shown in Figure 5C, the methane production rate reached a maximum of 0.30 ml CH₄/g VSS-hr at the 0.5 m height and declined rapidly to less than 0.1 ml CH₄/g VSS-hr in the upper levels of the unit. A similar pattern of activity versus filter height was reported by Van den Berg and Lentz (9). This trend in activity parallels the solids accumulation, COD, and volatile acids profiles reported previously, and verifies the trends in COD removal per unit of biological solids shown in Figure 5B. The variable activity suggests that the volatile solids in the upper

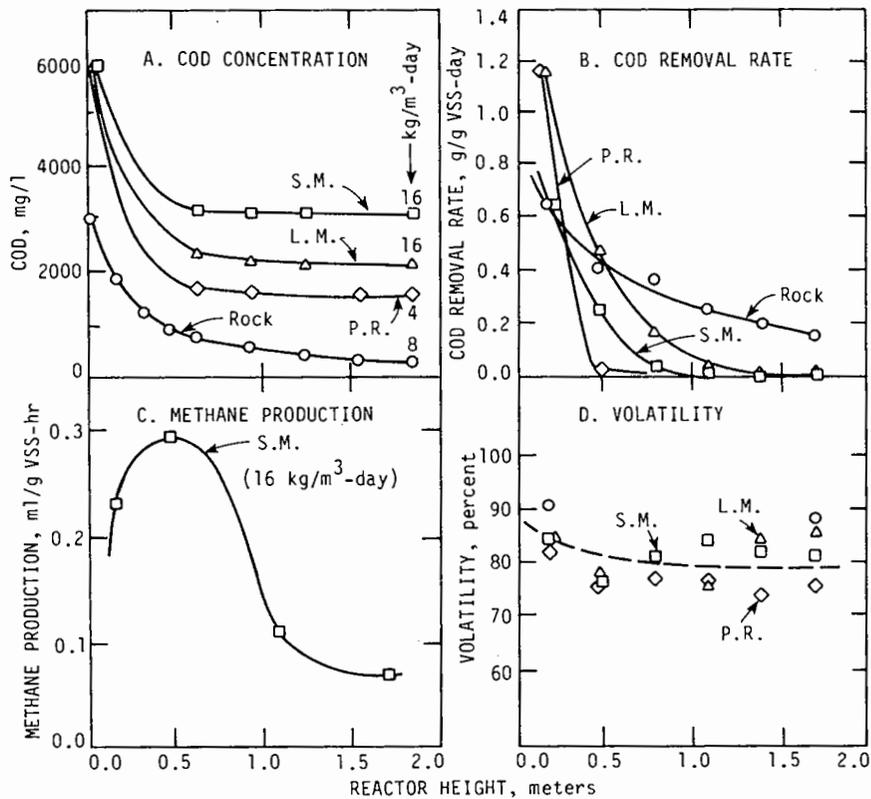


Figure 5. Activity and volatility of the biological solids distributed throughout the height of anaerobic filters.

sections of the reactor contained greater amounts of inactive cell matter than did the solids in lower sections. This lower activity in the upper levels also indicates that solids conditioning through biological decay was taking place.

Similar tests of activity could not be conducted for the attached solids because of the difficulty of removing the solids from the support medium without changing the

physical characteristics or exposing the microorganisms to oxygen.

The volatility of the suspended growth typically ranged from 75 to 85 percent throughout the height of the reactors (Figure 5D). There was a slight tendency toward higher volatility at the lower heights as might be expected in view of the results of activity measurements.

Visual observations and analytical tests indicated that units receiving wastes having high concentrations of carbohydrates contained a significant fraction of biological solids in the lower heights that were dispersed. These solids did not filter or settle readily. At intermediate heights, the biological solids were well-flocculated and would filter and could be settled easily. In the upper levels, most of the solids were attached to the media surfaces except for a small fraction of dispersed solids that seemed to migrate upward or downward depending on the extent of mixing or flotation caused by gas production and percolation through the medium (See Figure 4).

Visual inspection of the submerged media showed that the biological solids attached to the media were present in a layer of about 3 mm thickness with numerous places having large filaments extending into the liquid. These filaments collapsed when the media modules were removed from the liquid, leaving a film of biological solids of about 2 mm thickness.

Microscopic examination showed more clearly the physical nature of the biological solids. A significant part of the suspended growth consisted of well-flocculated or granulated matter. These granules had well-rounded edges and seemed to be bound together with filamentous forms of microorganisms. While it was difficult to associate performance with the size and appearance of biological solids, visual observations indicated that the onset of granulation of biological solids coincided with rapid improvements in COD removal and methane production. Therefore, media-related factors contributing to this granulation would be important considerations in selecting media for full-scale anaerobic filters.

DISCUSSION

The laboratory studies by Young (1, 2) and Dahab (7) and subsequent analyses summarized in this paper show that a number of factors affect the retention and distribution of

solids in fixed film, upflow anaerobic reactors. A number of these factors are media-related while others are related to hydraulic and other physical factors within the reactors. An assessment of the effect of these factors on biological solids retention and distribution and their importance to design and operation is given in the following discussion.

Biological Solids Accumulation, Distribution and Activity

Profiles of COD and suspended solids throughout the height of a number of laboratory-scale anaerobic filters show that most of the COD removal occurred in the lower 25 to 30 percent of the reactor height (See Figure 3). This removal was associated with a high concentration of biological solids that approached 60,000 mg/l in the lower sections of some reactors (See Figure 4). One-half to two-thirds of the biological solids mass in the lower one-third of the reactor height typically was not attached to the media matrix. Essentially all the biological solids in the upper one-third of the reactor height were attached to the media surfaces. Analysis of the activity of these solids showed that the solids at the lower heights were more active, as measured by their COD removal and methane production per unit weight, than were solids removed from upper heights of the reactor (See Figure 5).

This distribution of COD removal, biological solids accumulation and activity indicates that biological solids are largely synthesized in the lower levels where COD and solids concentrations are highest. A significant part of these synthesized solids is transported upward as a result of gas flow and hydraulic lifting. Without this transport, the concentration would soon become high enough to plug the reactor at the lower heights. For example, consider a reactor, 2 meters in height, in which no solids transport takes place when operating at an organic loading of 16 kg/m³-day. Further, assume that 30 percent COD removal takes place in the first 0.3 m height (See curve for large modular media in Figure 5A) and that the net biological solids yield in this volume is 0.05g VSS/g COD-day. (This yield is expected to be low at the lower reactor height and for the type of waste treated in the rock and plastic-media filters.) The COD removal in the first 0.3 m height would then be 9.6 kg/day and the net rate of production of biological solids would total 0.48 kg/day. In only 50 days of operation the accumulated solids concentration would be

80 kg/m³ or 80,000 mg/l. The measured concentration of biological solids did not approach this concentration in any of the reactors tested even when operated continuously for over one year, and the data shown in Figure 4D for the reactor containing the large modular medium suggests that a maximum concentration had been reached in the lower 0.3 m height of this reactor.

AS these solids reach the upper levels of the filter, the COD removal and gas production rate per unit mass of biological solids becomes quite low (See Figure 5B, C). This reduced activity suggests that in the upper reactor height, net growth is negative. That is, the biological solids carried up from lower heights are effective for accomplishing low unit rates of COD removal but decay exceeds synthesis and the mass of solids is reduced to inactive cell material.

Gas scouring in the upper levels of the media is expected to cause some sloughing of attached solids. These solids can flow upward and be lost from the reactor as effluent suspended solids. However, if in-bed flocculation takes place, these solids eventually may become flocculated sufficiently that they will settle downward. This is the expected mode of formation and direction of movement of the large granulated solids observed in all reactors.

The net effect of this transport of biological solids is that the reactor eventually can be expected to become filled with biological solids that are largely inactive. These solids must be wasted to prevent plugging the reactor or eventual appearance of high concentrations of suspended solids in the reactor effluent. Therefore, it is important that the reactor be designed and the media selected so that these excess solids can be removed by flushing or by gravity drainage.

Media Related Factors

Unit Surface Area

The unit surface area of the media did not seem to affect COD removal or solids distribution substantially. COD profiles shown in Figure 3 follow essentially the same trend for rock, which had a unit surface area of about 30 m²/m³, as for the plastic media, which had unit surface areas ranging from 98 to 138 m²/m³. In fact, the rock medium provided slightly better removal than did the plastic Pall rings, even though the rock media units were operated

at 25°C while the plastic media units were operated at 30°C. The two modular media provided almost identical COD removal at all reactor heights when operating at the 2 kg COD/m³-day loading, yet the unit surface area for the small-size material was over 40 percent greater than for the large-size.

The similarities in COD removal performance for media having such a wide range of unit surface areas is thought to have occurred because the majority of the biological reaction was associated with the loosely held solids and not with the attached solids. This conclusion is verified by the COD removal rate per unit of solids, which was essentially zero in the upper levels of the reactors where almost all the solids were attached.

Porosity and Pore Size

Little association could be made between performance and porosity. As shown in Figure 3, the overall COD removal performance in the rock medium, which had a porosity of 0.42, was essentially the same as in the plastic media having porosities of about 0.95. Greater differences were observed between the three plastic media.

The media-related factor seeming to have the most significant effect on COD removal performance and solids retention and distribution was the pore size. While pore configuration may have had some effect, the two modular media had essentially the same relative shape but the smaller medium had an average pore diameter of 32 mm as compared to 46 mm for the large modular medium. Yet the COD removal was substantially lower in the reactor containing the modular medium. The pore size in the Pall ring medium was estimated to be 20 mm, and the reactor containing this medium had even lower COD removal efficiency than did the reactor containing the small modular medium.

The lack of COD removal in the upper 1.5 meters of height in the reactor containing the Pall ring medium, as compared to low but measurable reduction in the upper levels of the other media (See Figures 3 and 4), suggests that greater short-circuiting was occurring in this medium. It was not possible from the tests conducted by the authors to credit this difference entirely to channeling or to differences in solids accumulation, but either factor could have caused the same differences in performance.

Summary

There is little doubt that the factors discussed above --excess biological solids accumulation, channeling, biological solids transport, suspended versus attached solids and media size, shape, porosity and pore size--all affect performance to some extent, and the interrelationship between these factors is expected to be quite complex. However, Young (2) and Dahab (7) succeeded in establishing reasonable agreement between measured COD removal performance and the performance simulated by use of a computerized model that integrated the effects of these factors with biological growth and substrate utilization kinetics. These modeling efforts, combined with measurements from laboratory-scale reactors operating over a broad range of operating conditions, helped to establish a basis for developing criteria for selecting media and for sizing full-scale anaerobic filters.

CONCLUSIONS

The laboratory-scale tests described above provide a basis for drawing the following conclusions:

1. Media type, size and shape are important factors to consider in designing full-scale anaerobic filters.
2. Unit surface area is not as important as a design parameter as is pore size and shape.
3. Biological solids retention and distribution is not affected substantially by media type, size or shape.
4. The majority of COD removal in fixed-bed, upflow anaerobic filters is associated with biological solids held loosely within the interstitial void spaces in the lower one-third of the reactor height.
5. The COD removal and methane production per unit of biological solids mass is substantially lower at upper reactor levels as compared to that at lower heights.

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APPLICATION OF STANDARD RATE AND HIGH RATE
ANAEROBIC TREATMENT PROCESSES*

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INTRODUCTION

In anaerobic biological treatment processes, organics are decomposed in a controlled, oxygen-free environment. Because oxygen is not required for decomposition, anaerobic processes are inherently one of the more energy-efficient means for stabilizing organics. As an additional benefit, they have the potential of capturing over 90 percent of the biodegradable energy contained in the wastewater organics in the form of methane gas, an easily transported, clean-burning fuel. When properly applied, anaerobic treatment can result in a net production of energy, as opposed to alternative stabilization methods, which are essentially all net consumers of energy.

Table 1 lists some of the benefits of anaerobic treatment. Important for many industrial treatment applications, anaerobic processes require substantially less nutrients (nitrogen and phosphorus) than their aerobic counterparts due to lower biomass yields. Nutrient addition can be reduced by a factor of five in some cases. However, low biomass yield is also the principal cause of several disadvantages of

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anaerobic treatment, which are listed in Table 1, as well, including a necessity for relatively long solids retention times (SRTs) to effect stable operation. SRT is a key factor in the application of anaerobic processes.

TABLE 1. ADVANTAGES AND DISADVANTAGES OF ANAEROBIC STABILIZATION PROCESSES (1)

Advantages	Disadvantages
<ul style="list-style-type: none"> ● High Degree of Stabilization ● Low Production of Waste Biological Sludge ● Low Nutrient Requirements ● Low Energy Requirements ● Methane Gas is Useful End Product 	<ul style="list-style-type: none"> ● Slow Growth Rate of Methanogens ● Requires Long Solids Retention Times ● Methanogen Sensitivity ● May Require Auxiliary Heating

In most cases, the advantages of anaerobic treatment far outweigh the disadvantages. Despite this fact, anaerobic processes are often not employed in applications where they are favorable, and many facilities that employ anaerobic treatment do not effectively use the energy-advantage of these processes to its full potential.

Although, in the past 50 years the fundamental understanding of anaerobic processes has been improved significantly through extensive laboratory- and pilot-scale studies (which has resulted in the development of several new processes), application of fundamentals to process design and use of the new anaerobic processes has been quite limited. The objective of this paper is to review the practical aspects of applying anaerobic processes for wastewater treatment, in an attempt to bridge the communication gap between research and practical application of new techniques. In order to accomplish this goal, the fundamentals governing anaerobic treatment are discussed initially, followed by a review of the three principal types of anaerobic reactors: conventional, anaerobic contact process, and submerged media anaerobic reactors (SMAR). It is important to note that each reactor type has specific advantages and disadvantages which dictate applications thereof. In recognition of this fact, a comparison of alternative techniques and applications is also presented.

ANAEROBIC TREATMENT FUNDAMENTALS

An understanding of the fundamentals of anaerobic decomposition is helpful for proper selection, design, and operation of anaerobic treatment systems. This is especially true when considering the relative merits and potential applications for new processes such as the anaerobic contact processes and SMARs. A brief review of the general principles of anaerobic treatment and the influence of environmental and operational factors is presented in the following relative to completely-mixed, suspended growth systems. The relationship of these fundamentals to each different design configuration is described in the respective sections.

Decomposition of organics in anaerobic processes is an extremely complex, symbiotic interaction of a variety of anaerobic and facultative bacteria. For simplicity, decomposition of complex organics can be characterized as a sequential, three-stage process comprised of: hydrolysis, organic acid formation, and methane fermentation. When the process is in balance, these three separate steps occur simultaneously at approximately equivalent rates.

There are numerous different types of bacteria involved in digestion, each characterized by their ability to use a relatively limited number of organic compounds as a food or energy source. Importantly, microbial growth rates and relative response to changes in the environment (reaction conditions) vary among groups. In general, the methanogens, that use low- and high-molecular weight fatty acids as a food source, have the slowest growth rate and are the most sensitive to environmental changes (5). Furthermore, methanogens are strict anaerobes that are inhibited by small amounts of oxygen (1).

The key factor in design and operation of anaerobic processes in general is to provide conditions conducive towards maintaining a large, stable population of methane bacteria. Typically, system inadequacies will be manifest in relatively high volatile acids concentrations in the reactor, i.e., imbalance between the populations of acid-forming bacteria and the methanogens.

Many factors are important in maintaining a stable population of methanogens, including: pH, alkalinity, temperature, nutrient availability, toxic nutrients, substrate characteristics (e.g., composition, biodegradability, etc.), and solids retention time. In this discussion, we will focus on solids retention time (SRT), which is certainly one of the

most important design and operational factors, recognizing the significance of other process variables.

For simplicity, SRT can be defined as the ratio of the total mass of microorganisms within a biological reactor with respect to the rate of loss of microorganisms for the same reactor. In other words:

$$\text{SRT} = M_R / \phi_e \dots \dots \dots \text{Eq. 1}$$

Where:

SRT = solids retention time

M_R = total biomass within the reactor

ϕ_e = biomass flux leaving the reactor = M_e/t

M_e = total biomass leaving the system including both that deliberately and that passing out with the effluent

t = time

More specifically, in the application anaerobic processes, we are interested in the SRT of the most sensitive microorganisms in the system, which are typically the slowest growing microbes -- the methanogens. We cannot define relative populations of organisms very accurately, and thus generally an alternate measure is typically used, such as total or volatile solids content. It is important to note that system performance is dictated by the rate limiting step in digestion, which is typically the conversion of fatty acids to methane as effected by methanogens, even though we may be forced to use a gross measure of the biopopulation of concern. Because methanogens are slow-growing microorganisms which are relatively sensitive to environmental changes, the key to maintaining effective, stable performance is the ability to maintain a long system SRT, i.e., a large population of active microorganisms. Also, the lower the operating temperature, the slower the growth of microorganisms -- hence, proportionately longer SRTs are required at lower temperatures to effect a similar degree of stabilization. For example, at ambient temperatures (10 to 25°C) an SRT in excess of 40 days is typically required for stable performance; whereas, in the mesophilic range (30 to 40°C), effective performance can generally be achieved at an SRT of 10 days or less. The different reactor configurations provide different environmental conditions for maintaining the anaerobic culture, thereby affecting the process SRT.

ANAEROBIC TREATMENT PROCESSES

There are four basic categories of anaerobic processes:

- conventional anaerobic digesters
- anaerobic contact processes
- submerged media anaerobic reactors (SMAR)
- anaerobic composting (semi-solid systems)

Anaerobic composting is applicable primarily to semi-solid materials such as biogasification of agricultural wastes (61). This is a special application which will not be covered herein.

Anaerobic processes are applicable for two purposes in wastewater treatment: (1) stabilizing organics with an associated potential for energy recovery, and (2) denitrification of wastewaters with high nitrate concentrations.

In conventional digesters, which are suspended-growth slurry reactors, the system SRT is approximately equal to the hydraulic retention time (HRT). Accordingly, relatively large reactors are required when processing dilute organic wastes at temperatures below 30°C, widely fluctuating or intermittent waste streams, or wastes with fluctuating toxicant loadings.

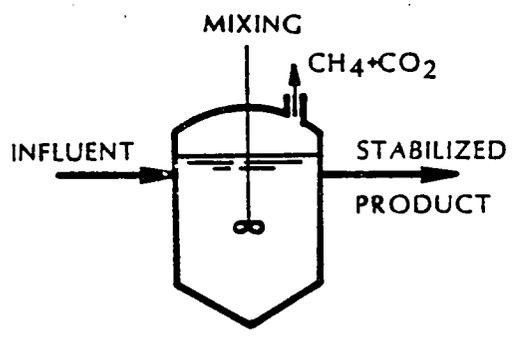
Anaerobic contact processes and submerged media anaerobic reactors (SMARs) were developed to accommodate more rigorous operating conditions, such as listed above. The basic principle of these processes is to preferentially retain active microorganisms in the reactor, particularly methanogens, thus increasing the system SRT for a given hydraulic loading. The following discussion briefly describes the major concepts of all three major reactor types, with emphasis on energy concerns. Denitrification with anaerobic processes will not be considered.

CONVENTIONAL ANAEROBIC PROCESSES

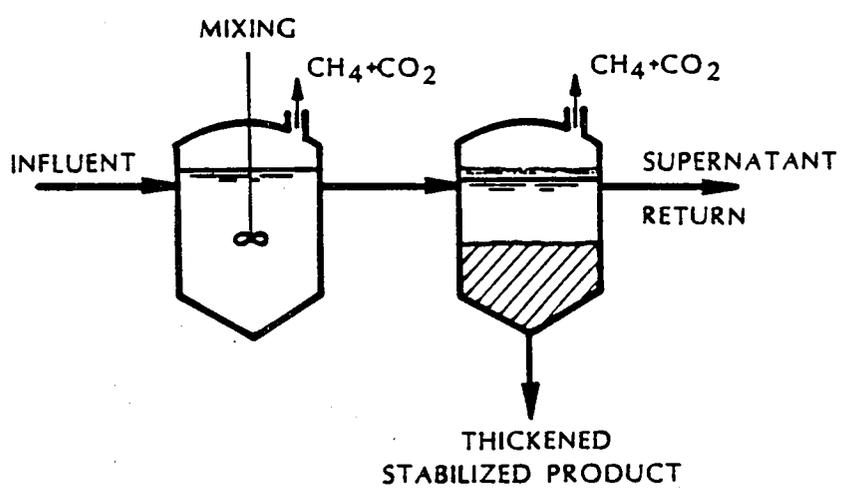
There are two basic configurations of conventional anaerobic digesters: 1) single-stage digestion, and 2) two-stage digestion, as illustrated in Figure 1. Also, single-stage digesters are typically categorized as either standard-rate (low-rate), or high-rate digesters.

Standard-Rate Digestion

Standard-rate digesters are single-stage reactors that are not mixed by supplemental means. Some degree of agitation



SINGLE- STAGE CONVENTIONAL DIGESTER



TWO-STAGE CONVENTIONAL DIGESTER

Figure 1. Schematic diagrams of conventional aerobic digesters.

is provided by natural gas evolution during digestion, convection currents caused by temperature gradients, and pumped recirculation for digester heating. These effects have been shown to be quite significant (17). Even so, it has been estimated that less than 50 percent of the volume is effectively utilized in conventional standard-rate digesters (18), although no definitive studies have been conducted to confirm this hypothesis. Conventional, standard-rate digesters are typically designed for HRTs (SRT) in excess of 30-days (30 to 60 days) and corresponding organic loadings up to 0.1 lb VS/day/cu ft (12).

High-Rate Digestion

Supplemental mixing is used in high-rate digestion in order to increase the "effective" digester volume thus permitting higher loading rates. Hydraulic loadings to high-rate digesters typically range from 6 to 30 days, corresponding to organic loadings as high as 0.5 lb VS/day/cu ft (18,19,20). Studies conducted in the early 1900's demonstrated that mixing of digesters increased the rate of stabilization substantially (21). However, high-rate digesters were not considered seriously in wastewater plant design until the 1950's when the relationship between HRT (i.e., SRT) and digester performance was clearly established (12). Prior to this time, system designs were based primarily on organic loading, e.g., lb VS/day/cu ft of reactor. When HRT was recognized as the fundamental design parameter, it became apparent that thickening the feed sludge to digestion would result in a more efficient utilization of reactor volume for an equivalent degree of stabilization. However, the importance of mixing is amplified at these higher feed solids concentrations, since the associated higher viscosity has a tendency to decrease the mixing effects of natural gas evolution and convectional currents (12), as well as reducing the effects of external mixers.

Two-Stage Digestion

The first stage (primary digester) of a two-stage system is generally a mixed, high-rate digester; whereas, the second stage (secondary digester) is an unmixed unit. The purpose of the secondary digester is to store and concentrate the product prior to ultimate disposal. In many applications, good separation and thickening does not occur (digestion of activated sludge in particular), thus negating the purpose of the

two-stage approach. The primary digester is generally designed using the principles of high-rate digestion.

Conventional Digester Process Considerations

The major process considerations for conventional digesters are heating and mixing. Supplemental heating is generally required to effect stabilization in a reasonable reactor volume.

Energy Requirements for Sludge Heaters and Recirculation Pumping

Digester heating is generally provided with an external, boiler-heat exchanger, fueled either by digester gas or natural gas. Other, less popular methods are sometimes used, e.g., direct stream injection, and hot water pipes inside a digester. The efficiency of a sludge heater typically ranges from 70 to 85 percent. Higher efficiencies are associated with new equipment which is well maintained. Heat exchanger scaling decreases efficiency significantly. In addition, electrical energy is required for pumped recirculation and miscellaneous water and oil pumps. Typically, approximately 6 to 10 hp is required per million Btu/h heater capacity, corresponding to an electrical energy demand of approximately 6 kW/million Btu heat transferred. Total energy requirements for sludge heating in the northern U.S. are summarized in Table 2 (28).

TABLE 2. TOTAL ENERGY REQUIREMENTS FOR SLUDGE HEATING OF COMPLETELY-MIXED DIGESTERS IN THE NORTHERN U.S.† (28)

Hydraulic Retention Time, days	Fuel for Heating, Btu/gal	Electrical Energy, kWh/1,000 gal	Total Energy*, Btu/gal
10	575	3.45	611
15	625	3.91	666
30	780	4.68	830
50	1,000	6.00	1,060

†Energy requirement referenced per unit volume processed, gallons

*Electrical energy conversion efficiency of 10,500 Btu/kWh.

Mixing in Anaerobic Digesters

The purpose of mixing in digestion is to (17):

- distribute organics, (i.e., food sources) within the reactor,
- prevent stratification (scum blankets and excessive sedimentation),
- distribute inhibitory substances within reactor,
- prevent temperature gradients, and
- permit effective utilization of digester volume.

Anaerobic Digester Mixer Types. Digester mixing is accomplished by "pumping" or recirculating the reactor contents. This is typically done with gas recirculation, liquid recirculation using external pumps, or internal, impeller mixers.

Gas mixers transmit energy and flow by gas-lift pumping, and thus mixer design is analogous to gas-lift pump design. The various gas mixers used include: unconfined gas spargers, draft-tube gas agitators, and piston gas mixers. Generally, multiple gas distribution points are required for good circulation, and certain designs use an automatic control system to sequentially direct gas to various locations in a reactor. The primary advantage of gas mixing systems is the lack of moving parts within the digester. Due to this fact, gas agitation is the most widely used method of mixing in digester applications.

In this method, sludge pumps are used to recirculate the digester contents through a controlled distribution system. Many digestion systems use recirculation pumps as an integral part of the heating system, and this recirculation supplements natural mixing and other mixers that may be employed. There are very few treatment plants that use pumped recirculation as the primary source of mixing, although the concept is simple and energy transfer efficiencies are relatively high. For effective mixing, this approach requires a rather extensive distribution system, similar to gas-recirculation design.

There are two types of impeller mixers employed: (a) high-speed, propeller mixers, and (b) low-speed, pitched-blade turbines. Propeller diameters ranging from 1 to 2 feet are used for high-speed mixers, operating at speeds from 300 to 600 rpm. Rotating speeds of low-speed mixers are substantially less, ranging anywhere from 15 to 30 rpm, and corresponding impeller diameters are much larger, ranging from 5 to 15 feet.

High-speed propeller mixers have been used extensively for mixing digesters; however, inefficient operation and high maintenance are unfavorable characteristics of this approach. Low-speed turbines have not been widely used in digester mixing, possibly because of the poor track record of propeller mixers. Low-speed turbines, however, are the most efficient devices available for mixing digesters, and with the trend towards energy-efficient design, it is likely that this approach will be used more frequently in future designs.

Experience in Digester Mixing. Although the benefits of mixing in anaerobic digestion were recognized in the early 1900's, to be later confirmed through laboratory and pilot-scale testing in the 1950's and 1960's, there were no field tests conducted on mixing in full-scale digesters until the mid-1970's.

Several recent studies have evaluated mixing in full-scale digesters through dye- and radioactive tracer analyses (17,19,20,21). In general, these studies have demonstrated that the present state-of-the-art in mixer design is not very effective.

The findings of a recent comprehensive mixing study by Smart (30) under the auspices of the Ontario Ministry of the Environment were quite alarming. This study evaluated mixing efficiency of 10 selected digester facilities using dye-tracer tests. Facilities were selected to include representative mixer types and a wide range of nameplate power intensities: six gas recirculation mixers and four high-speed propeller mixers were evaluated. The nameplate power intensities of the gas recirculation mixers ranged from 0.04 to 0.2 hp/1,000 cu ft of reactor volume, with corresponding gas recirculation rates varying from 0.3 to 3 CFM/1,000 cu ft. Nameplate power intensities of the propeller mixers ranged from 0.2 to 0.5 hp/1,000 cu ft. Digester sizes ranged from 40- to 100-foot diameter, and theoretical HRT's were from 10- to 45-days at the flow rates used during testing.

The major conclusion of this study was that, "mixing of the anaerobic digesters was, in general, grossly inadequate with respect to volume utilization." Essentially all of the digesters exhibited extensive short circuiting and poor volume utilization (i.e., extensive dead zones). In general, less than 50 percent of the actual volume represented the "effective" digester volume, and therefore the actual system HRT was less than one-half the theoretical HRT. In addition, the researchers were unable to identify any meaningful

relationship between nameplate horsepower and mixing intensity, or any clear-cut difference between gas-recirculation mixers and high-speed, impeller mixers. They recommended additional full-scale studies for optimization of mixing and process design.

Recent Advances in High-Rate Digestion - Improved Mixing

A recent study of mixer applications in the chemical process industry (28) showed that for most large-tank applications, slow-speed impeller mixers are inherently more efficient agitation devices than gas-agitation mixers. However, as with gas mixing, recirculation efficiency is a function of system design. There are two basic impeller mixers available:

- high-speed, propeller mixers, and
- low-speed, axial turbines.

The effect of a mixer is directly related to its pumping capacity, which for turbulent flow impellers is described by the following (25):

$$Q = N_q \cdot ND_a^3$$

Furthermore, the power delivered to the liquid is directly related to discharge according to:

$$P = \rho QH$$

Where, Q = impeller discharge

N_q = impeller discharge coefficient which is a function of its configuration

N = rotational speed

D_a = impeller diameter

P = power

ρ = fluid density

H = discharge head

Thus, for a given impeller type, discharge is directly related to impeller speed, N, and is proportional to the third-power of impeller diameter, D_a . The significance there is that large-diameter, slow-speed turbines are more efficient than high-speed, propeller agitators: in addition, high-speed impellers are subject to more maintenance problems than low-speed units. Despite these facts, most mechanical digester mixers are of the high-speed, propeller type. Recently, the City of Las Vegas, Nevada, applied mixing technology used in the chemical processing industry for improving digester volume utilization. They have installed five, low-speed, axial-turbine mixers designed to mix over 80 percent

of the actual liquid volume. A description of this conversion follows.

In order to increase the capacity of the digestion facilities at the City of Las Vegas, five of the six existing 66-foot diameter digesters were retrofitted with low-speed, axial turbine agitators to provide more positive mixing at a substantially higher mixing intensity. The purpose of this change was to increase the "effective" digester volume and hydraulic retention time, thus, improving the safety factor of design. Table 3 summarizes the mixer design for each unit.

TABLE 3. SUMMARY OF AXIAL-TURBINE MIXER DESIGN FOR LAS VEGAS, NEVADA

Digester Dimensions	66 ft diameter x 28 ft swd
Volume	95,750 cu ft
Mixer Type	Twin impeller, axial-flow turbine
Nameplate Horsepower	60 hp
Delivered Water Horsepower	48 hp
Rotative Speed	25 rpm
Circulation Rate	300,000 gpm
Superficial Velocity	11 ft/min
Upper Impeller Design	
Diameter	98 inches
Pitch	32°
Circulation Rate	75,000 gpm
Lower Impeller Design	
Diameter	124 inches
Pitch	32°
Circulation Rate	225,000 gpm

Axial-flow turbines transmit power through large, low-speed (15 to 30 rpm), pitched-bladed turbines, inclined on an angle 32° or 45°. Baffling is generally required to assure uniform distribution and to prevent rotation of the entire digester contents.

At the time of this writing, the axial-turbine mixers in Las Vegas had been operating for approximately 9 months. During this period, digester performance has been good. Although a comprehensive study has not been conducted to properly characterize the effect of more intensive mixing on system performance, several subjective observations indicate operation has been improved significantly from the past

conventional gas-mixing operation. For example, one digester has been loaded to a HRT of 8 days for extended periods of time without adverse operational effects. In contrast, the digester that was not renovated and still employs gas mixing, was unable to operate successfully at this high loading. Also, plant operators have observed an accumulation of trickling filter snails in the overflow box (never before observed), which is indicative of relatively high mixing intensity to keep these relatively heavy objects suspended. And, as a final note, in contrast to high-speed impeller mixers, ragging has not proven to be a problem even though 24-hour-per-day operations are maintained.

Summary of Total Energy Requirements for High-Rate Digesters

For comparison with alternative high-rate digestion processes, Table 4 summarizes the energy requirements for conventional digesters. Heating comprises the major energy demand, ranging from 60 to 80 percent of the total consumption. However, these data do not accurately represent energy for mixing at long SRTs, since as SRT is increased, the importance of uniform mixing is lessened. In fact, at a 50-day SRT, in most cases, little or no mixing is required. Importantly, the data indicate the energy "break even" point for conventional digestion. Assuming 80 percent recovery and utilization of the digester gas produced and 90 percent stabilization, then an anaerobic digester operating at a 15-day SRT will be energy self-sufficient if the influent BOD_L is above 28,000 mg/L. The importance of concentrating the feed to conventional digesters and good mixing is apparent.

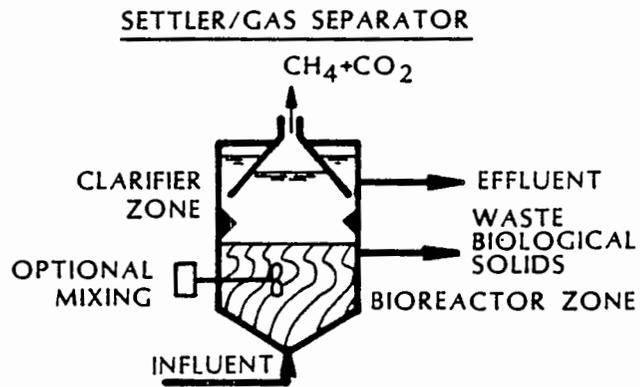
ANAEROBIC CONTACT PROCESSES

The anaerobic contact process was developed to overcome some of the problems of conventional digestion; in particular, the requirement of long hydraulic retention times, (i.e., large reactor volumes) to achieve adequate SRTs for stable performance. The anaerobic contact process is similar to the activated sludge process (Figure 2). It is comprised of a completely-mixed slurry reactor and a settling basin (or zone) in series. Solids and microorganisms that are separated in the settling basin are recycled to the anaerobic reactor. In this way, it is possible to retain a relatively high inventory of active microorganisms at high HRTs, i.e., the SRT is greater than the HRT. Accordingly, a much smaller

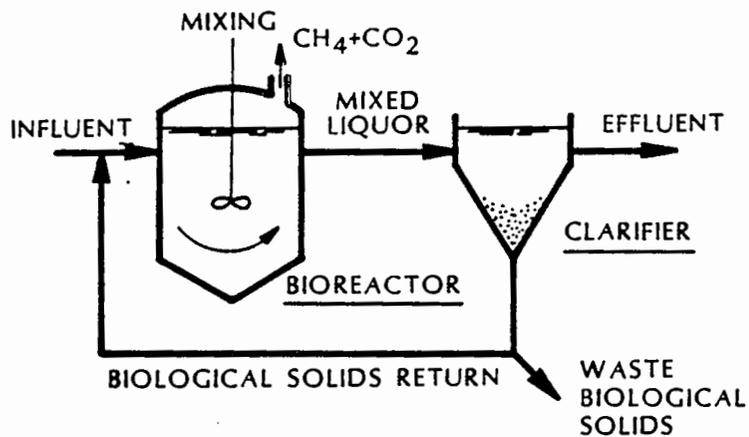
TABLE 4. TOTAL ENERGY REQUIREMENTS FOR COMPLETELY-MIXED,
CONVENTIONAL DIGESTERS*

Hydraulic Retention Time, days	Heating		Mixing, kWh/gal	Total Energy			Heating % of Total
	Fuel, Btu/gal	Electrical, kWh/gal		Fuel, Btu/gal	Electrical, kWh/gal	Combined, Btu/gal	
10	575	0.003	0.012	575	0.015	736	83
15	625	0.004	0.018	625	0.022	856	78
30	780	0.005	0.036	780	0.041	1,210	69
50	1,000	0.006	0.060	1,000	0.066	1,690	63

*Energy requirement referenced per unit volume processed, gallons.



UPFLOW ANAEROBIC CONTACT PROCESS



ANAEROBIC ACTIVATED SLUDGE PROCESS

Figure 2. Schematic diagram of anaerobic contact process.

reactor is possible for a given degree of stabilization and reliability. Two reactor configurations are typically employed, an upflow reactor and an anaerobic activated sludge system.

As with activated sludge, solids settleability and solids separation efficiency are the keys to effective treatment. Lettinga, et al (26) have found that a "maturing" period is required to develop granules, believed to be bacterial growth, which tend to settle well and remain in the system. Stable operation occurs when this granular growth develops. Even under optimal operating conditions, the organic content of the effluent is high with respect to secondary treatment standards. Effluent BOD concentrations are typically 200 to 1,000 mg/L, depending on the influent waste strength and loadings. Nevertheless, the anaerobic contact process is an extremely effective and energy-efficient pretreatment process, which can be used for stabilizing relatively high-strength industrial wastes prior to subsequent polishing treatment. Some example operations of the anaerobic contact process in this capacity are shown in Table 5.

The results shown in Table 5 illustrate the potential of the anaerobic contact process as a pretreatment method. In these applications, most of the operating temperatures were in the mesophilic range, above 30°C, except the potato waste trials at reduced temperatures. At this time there is insufficient evidence to predict performance at suboptimal temperatures (less than 30°C), but in the mesophilic range, performance was obviously quite good. Stabilization efficiencies range from 65 to over 90 percent, typically 85 to 90 percent, and at relatively high organic loadings. Importantly, the process performed well at relatively low hydraulic loadings (0.5 to 7 days) where conventional digestion would be ineffective. Thus, the concept of increasing the SRT by solids recycle is valid for anaerobic treatment as well as for aerobic processes. In accordance with a smaller reactor volume, the heating requirements would be reduced due to a reduction in heat loss to surroundings.

Presently, there is insufficient information for optimizing design of the anaerobic contact process. Pilot studies would be required for effective utilization of this concept in most applications. However, certain guidelines have been proposed (26), as indicated in Table 6.

The predominant energy requirement for the anaerobic contact process is fuel for heating the influent. Optimal operating temperatures have not been established as yet.

TABLE 5. PERFORMANCE SUMMARY FOR SEVERAL ANAEROBIC CONTACT PROCESS ALTERNATIVES (2,26)

Waste Type	Temperature, °C	Hydraulic Retention Time, Days	Influent		Organic Loading lb/d/cu ft		Treatment Efficiency, %	Reference
			COD, mg/L	BOD, mg/L	COD	BOD		
Brewery	-	2.3	-	3,900	-	0.13	96	2
Citrus	33	1.3	-	4,600	-	0.21	87	2
Gotton Kiering	30	1.3	-	1,600	-	0.07	67	2
Molasses	33	3.8	-	32,800	-	0.55	69	2
Meat-Packing	33	1.3	-	2,000	-	0.11	95	2
Meat-Packing	33	0.5	-	1,380	-	0.16	91	2
Meat-Packing	33	0.5	-	1,430	-	0.16	95	2
Meat-Packing	33	0.5	-	1,310	-	0.15	94	2
Meat-Packing	33	0.5	-	1,110	-	0.13	91	2
Potato	19	1.2	3,000	-	0.25	-	92	26
Potato	26	3	3,000	-	0.78	-	89	26
Potato	30	4	3,000	-	1.00	-	89	26
Potato	35	6.5	3,000	-	1.87	-	89	26
Sugar	30	4	5,000	-	1.50	-	92	26
Sugar Beet	30	5	3,500	-	1.94	-	80	26
Starch-Gluten	35	3.8	-	14,000	-	0.10	80	2
Wine	33	2.0	-	23,400	-	0.73	85	2
Yeast	33	2.0	-	11,900	-	0.37	65	2

TABLE 6. GUIDELINES FOR APPLICATION OF THE ANAEROBIC CONTACT PROCESS (26)

● Settling Velocities	<2.3 ft/h
● Organic Loadings	<2.5 lb COD/d/cu ft
● Hydraulic Loadings	>3 hours
● Upflow Reactor Distribution Spacing Mixing	One per 100 sq ft of bottom area
● Waste Strength	>1,000 mg/L BOD _L

Based on field data, it appears that a temperature of 30°C (86°F) would be adequate in most cases. This corresponds to an energy requirement of 300 Btu/gal, assuming a 50°F influent temperature (industrial waste temperatures vary considerably depending on processing technology). Assuming a capture efficiency of 80 percent of the digester gas and 90 percent stabilization efficiency, the anaerobic contact process will be energy self-sufficient if the influent BOD_L concentration is above 9,800 mg/L, as compared to 28,000 mg/L for conventional digestion.

SUBMERGED MEDIA ANAEROBIC REACTOR (SMAR)

As mentioned, conventional anaerobic processes are limited primarily to the treatment of high-strength wastes, such as municipal waste sludge, due to long hydraulic detention times necessary to achieve stable operation (i.e., long SRTs). Although the anaerobic contact process is effective for treating wastes containing more than 1,500 mg/L BOD_L, it is limited primarily to pretreatment applications because of limitations in separating effluent solids cost-effectively, and because it is not applicable to widely fluctuating or intermittent loadings. Also, when treating low-strength wastes, methane is not generated in sufficient quantities to heat the waste stream, as may be required for cost-effective and energy-efficient treatment. During the past decade, the submerged media anaerobic reactor (SMAR) was developed to accommodate more rigorous operating conditions than possible with conventional anaerobic processes and the anaerobic contact process (3).

Process Description and Design Configurations of the SMAR

The SMAR employs an inert support medium to retain biomass within the reactor, thus developing long SRTs which are

necessary for effective anaerobic treatment. As such, SMARs are capable of effective treatment at relatively short hydraulic detention times in reactors of simple design. The retention of a large biomass inventory in the reactor allows operation at nominal temperatures and lends considerable stability to the process. The SMAR system resists transient conditions and system stresses quite well.

Depending on reactor design, the SMAR can be applied either as a roughing step for pretreatment of high-strength wastes, or as a secondary treatment process to produce a high-quality effluent. There are two major design configurations for the SMAR; the static bed reactor and the expanded bed reactor. For reference, the static bed SMAR was initially referred to as the "anaerobic filter" and the expanded bed SMAR as the "anaerobic attached film expanded bed reactor" (AAFEB). The term SMAR was recently adopted by an ad hoc committee on Anaerobic Fixed Film Bioreactors, to alleviate some of the ambiguity of previous nomenclature (27). A brief description of these two reactor types follows.

Static Bed SMAR

The basic design configuration for the static bed SMAR is illustrated schematically in Figure 3. In this process, the waste stream is introduced at the bottom of the reactor and passes upward through a support medium (either rocks or synthetic media similar to that employed in trickling filters). The treated effluent and biogas are separated and removed at the top of the reactor. Biological solids adhere to the surface of the support media and flocculate in the interstitial spaces establishing a large biomass inventory.

The static bed SMAR functions as a bacterial film anaerobic process, although suspended growth organisms and granules (similar to anaerobic contact) are also important. Stabilization of waste takes place primarily at the surface layer of biological solids. The hydraulic regime approaches a plug flow mode in the reactor, although the upward flow of gas through the column causes significant deviation from "ideal" plug flow. The combination of anaerobic bacterial film process with a plug flow reactor configuration, results in a stable and efficient biological process, indicating that methanogens are retained quite well within the reactor. High rates of stabilization are achieved in the lower sections of the reactor where both substrate concentration and biological activity are high. As the waste flows through successive

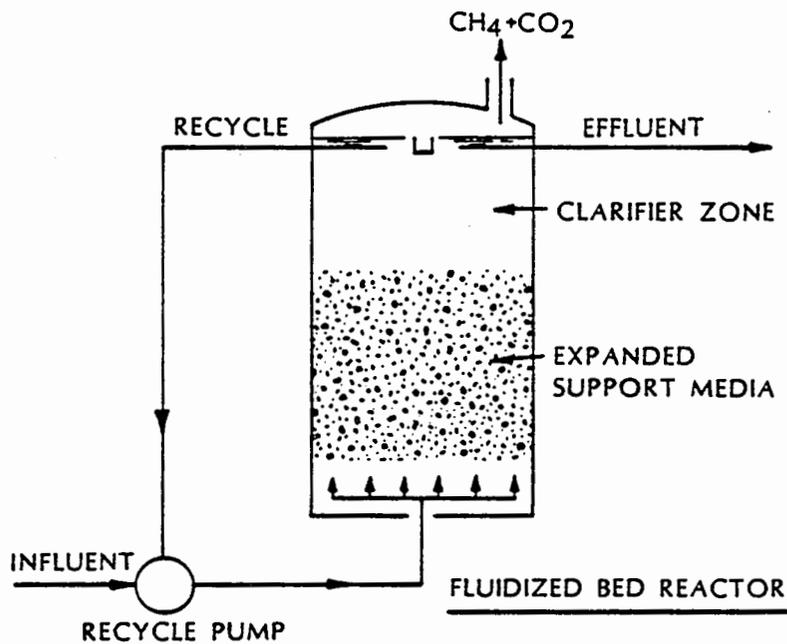
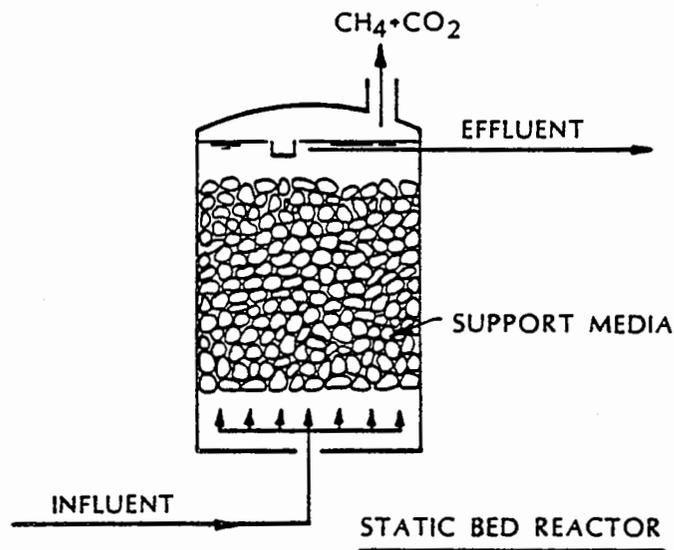


Figure 3. SMAR design configurations.

layers of the reactor, organic material is removed continually by the active biological solids within that layer. Cellular materials produced in the lower levels pass upward through the filter by the combined effects of liquid movement and gas flow. The effluent is well stabilized, containing little biodegradable material and effluent solids concentrations are typically low.

The mode of operation of the static bed SMAR can be likened to that of a number of reactors in series, where organic removal rates are high in the initial sections, and polishing and solids separation occur in subsequent areas of the reactor. The major advantages of this process are: (1) an effluent can be produced having much lower concentrations of unstabilized organics than is possible in a single-stage system, and (2) environmental conditions in each stage may favor the development of waste-specific microorganisms which could increase the rate of consumption of a particular component in a waste.

Expanded Bed SMAR

The second type of submerged media anaerobic reactor to receive considerable attention is the expanded bed SMAR. The basic design configuration for this process (Figure 3) consists of a column packed with a bed of small, inert particles 0.5 to 1 mm in diameter. The bed is expanded slightly by the upward flow of the waste stream through the column. In order to avoid high flow rates, the expanded bed reactors are operated just above incipient fluidization rates (5 to 20 percent bed expansion). This minimizes energy input, yet is sufficient to ensure good contact between the biofilm and the liquid wastes. Bed expansion is usually achieved by closed-loop recirculation of a portion of the effluent, thus enabling constant fluidization velocities under variant system hydraulic loadings. To further reduce process energy inputs, a low-density support media is employed; however, media density must be kept high enough so that it will be retained in the reactor.

The use of the small inert media provides a large surface area for the development of microbial film. When used in the expanded bed mode, this system permits the development of long SRTs and a dense and concentrated biomass, while maintaining thin biofilms. As such, this process overcomes several shortcomings of static media reactors, e.g., possible clogging and short-circuiting. Additionally, the use of

recycle creates a uniform distribution of solids throughout the bed.

In the expanded bed SMAR, waste is mixed with the recirculated portion of the effluent and passes in an upflow manner through the bed. The liquid effluent and biogas are separated and removed at the top of the reactor. Stabilization is achieved through contact with the biomass, where substrate removal efficiency is governed by biofilm removal rates. In addition, support media entrap small particulates, thus acting as a physical filtration device, further assisting in maintaining high biomass inventories at high HRTs.

Process Evaluation

Due to differences in design configurations, hydraulic regimes, and the nature of the retained biomass, static bed SMARs and fluid bed SMARs exhibit considerable differences in treatment performance and applicability. Therefore, each system is evaluated separately in the following.

Static Bed SMAR

Hydraulic detention time is the most significant operating parameter governing static bed SMAR treatment efficiency. Results from several studies (11,13,16,20) treating a wide variety of wastes have shown that removal efficiencies are quite high at relatively short hydraulic detention times, but decrease as the hydraulic detention time is decreased, as shown in Figure 4.

Several important observations can be made based on these data. For example, at constant flow, the "steady-state" effluent BOD_L concentration has been shown to vary directly with changes in influent BOD_L concentration; and conversely, at a given influent BOD_L concentration, the effluent quality varies in proportion to the flow rate. Treatment performance is also affected by reactor height; as shown in Figure 5, for a given hydraulic retention time, higher removal rates are achieved in shorter reactors.

Most importantly, static bed SMARs are able to achieve high removal efficiencies at organic loadings which are comparable to or higher than typical loadings applied to other biological processes. The success of this technology stems from its ability to retain biomass within the reactor for stabilization of the waste stream. The large mass of biofilm provides effective organic removal, and compensates for

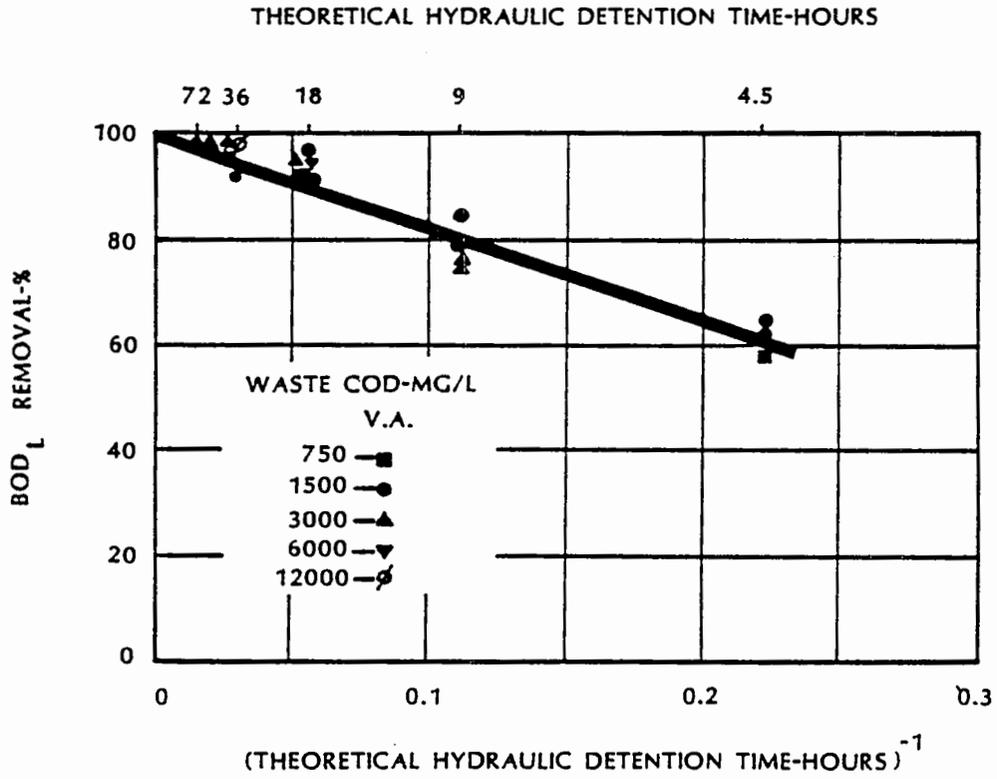


Figure 4. Treatment efficiency versus the reciprocal theoretical hydraulic detention time when treating the volatile acids and protein carbohydrate wastes at strengths ranging from 750 to 12,000 mg/L as COD (3).

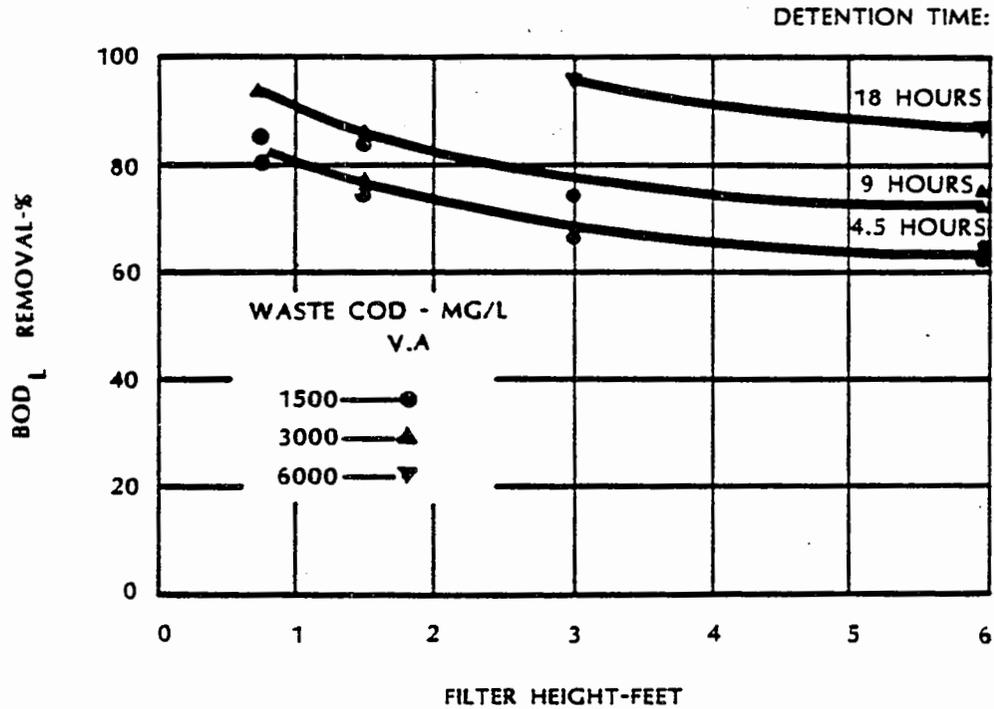


Figure 5. Treatment efficiency versus filter height when treating the volatile acids and protein-carbohydrate waste at organic loadings of 0.212 lb COD/d/cu ft and waste strengths of 1,500 to 6,000 mg/L as COD (3).

decreased unit activity of microorganisms caused by adverse conditions and system stresses without permanent loss of treatment efficiency. (Additionally, with proper acclimation, SMARs perform effectively at non-optimal environmental conditions such as depressed pH, low temperatures, and in the presence of toxicants). Table 7 summarizes several important design and performance criteria for static bed SMARs.

TABLE 7. SUMMARY OF DESIGN AND PERFORMANCE CRITERIA FOR STATIC BED SMARS

Parameter	Comments
System Stability	Excellent, effective at non-optimal conditions (low pH presence of toxicants)
Loadings	Typically 0.20 to 0.50 lb BOD _L /d/cu ft
Load Variations	Up to 4-fold without upset (1.5-2 lb BOD _L /d/cu ft)
Intermittent Operation	Extended downtime - 6 to 9 months water rapid startup (matter of days)
pH	5.4 to 9.3
Temperature	20 to 35°C
Toxicants	Tolerate with acclimation excellent resilience
Media	High specific surface area void volume
Reactor Height	Shallow depth preferred, 6 to 15 ft
Recycle	Typically unnecessary

Static bed SMARs can perform well at normal ambient temperatures of 50°F or higher provided the influent BEP concentration is sufficiently high (greater than 2,000 mg/L). Therefore, supplemental heating is typically not required. Hence, in this process the principal energy demand is for influent pumping. The pumping head will vary depending on the applied loading. Typically the total dynamic head will be 30 feet or less. The electrical power requirement is therefore about 0.16 kWh/1,000 gallons, or less than 2 Btu/gallons treated.

The principal application of the static bed SMAR is as a pretreatment process for reducing high concentrations of organic material from soluble, biodegradable waste streams to manageable levels. The SMAR's ability to handle large organic loadings at relatively short hydraulic retention times while producing low quantities of sludge makes it ideally suited to such applications. Furthermore, the SMAR process is applicable to a wide variety of soluble wastes ranging from food processing and pharmaceutical wastes, to heat treatment liquors resulting from thermal conditioning of municipal sludge. The excellent ability of the SMAR for handling intermittent loads also makes it acceptable for treatment of wastes from industries operating only a few days each week, as well as for seasonal waste streams.

The static bed SMAR is primarily limited to treatment of soluble wastes, although small amounts of degradable suspended solids could be accommodated. The process exhibits poor performance when treatment waste strengths are below about 1,000 mg/L COD. As such, it is not attractive for treatment of domestic sewage.

As yet, only two full-scale applications of this process have been operated, although additional reactors are under design or construction at present. Accordingly, sufficient scale-up information for an optimized design are not currently available.

Expanded Bed SMAR

The success of the expanded bed SMAR in producing high quality effluents is due to the use of high specific surface area media, which enables the development of a sufficient biomass inventory for effective organic removal even at low substrate concentrations and creates a vast area for contact of the biofilm with the waste stream. The hydraulic expansion of the bed increases the effective surface area, and minimizes the potential clogging and short-circuiting problems associated with packed beds (3).

When treating dilute wastes (<600 mg/L BOD) such as primary settled sewage, a relatively thin biofilm develops (<15 micron). In this case, the kinetics of organic removal are controlled primarily by the rate of biological metabolism, and mass transfer limitations are relatively insignificant since full substrate penetration occurs throughout the entire film thickness (3). Therefore, organic removal kinetics can be described as similar to suspended growth systems.

Most data available to date indicate that effluent quality is a function of organic loading rate, as seen for several lab-scale work treatment applications in Figure 6. The results presented in Figure 6 show that effluent quality is directly dependent on both influent concentration and organic loading rate. Additionally, these data demonstrate that, at a given organic loading, higher removal efficiencies are achieved with stronger wastes, as might be expected.

The data presented in Figure 6 were derived using different wastes with varying non-biodegradable fractions, different support media, and different operating temperatures. Hence, a general empirical formula predicting effluent quality cannot be derived. Nonetheless, the available data are useful for indicating loading ranges conducive to a desired level of treatment. For example, if the treatment objective is to remove large quantities of organics, as in pretreatment of a high strength waste, then organic loadings in the range of 1.0 to 1.25 lb BOD_L/day/cu ft are appropriate. However, if the goal is a high quality effluent, such as meeting secondary treatment standards, then much lower loadings are required, in the range of 0.1 to 0.3 lb BOD_L/day/cu ft. Pilot studies would be required in most cases to verify performance estimates and many practical considerations, and applying expanded bed technology must be addressed for performance to be successful and reliable.

In contrast to the static bed SMAR, the expanded bed SMAR can effectively remove organics from dilute wastewaters while operating at short hydraulic detention times (several hours) and at operating temperatures as low as 10° to 20°C. Under these conditions, the process is capable of high organic removal efficiencies (60 to 90 percent) when treating low strength (200 to 600 mg/L COD), soluble wastes at organic loading rates of up to 0.50 lb/day/cu ft. This difference is due to the large specific surface area available using small support media in the expanded bed mode, which enables the development of larger biomass concentrations and higher solids retention times per unit volume of reactor. Table 8 summarizes pertinent design and performance relationships for expanded bed SMARs.

Expanded Bed SMAR for Secondary Treatment

There has been considerable interest recently in the possibility of using the expanded bed SMAR in lieu of conventional secondary treatment.

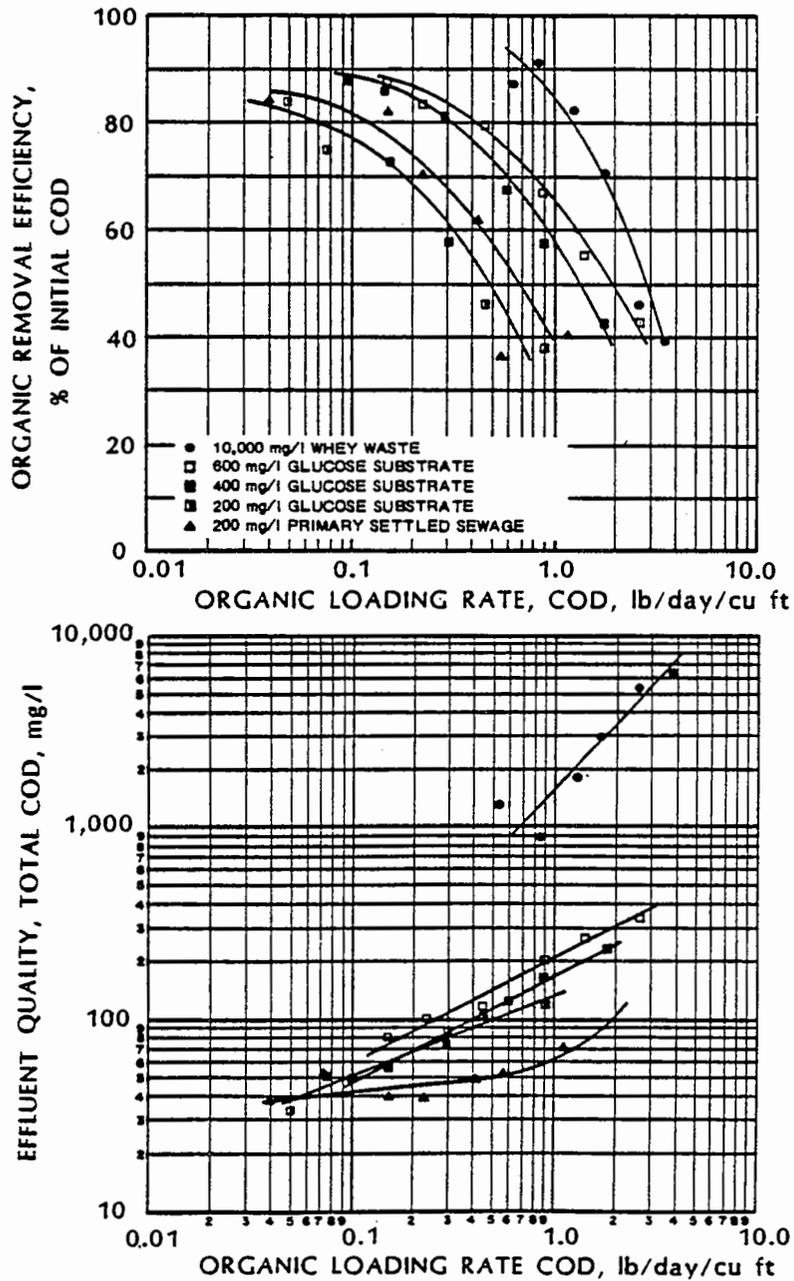


Figure 6. Organic removal efficiency and effluent quality versus organic loading rates in the expanded bed SMAR (3).

TABLE 8. SUMMARY OF DESIGN AND PERFORMANCE CRITERIA FOR EXPANDED BED SMARS

Parameter	Comments
System Stability	Excellent
Loadings	Pretreatment: 1-1.5 lb BOD _L /d/cu ft Secondary effluent: 0.1 - 0.3 lb BOD _L /d/cu ft
Load Variations	Excellent resilience, up to 10-fold increase. ERT as low as 5 minutes
Intermittent Operation	Assumed same as static bed reactor, excellent
pH	Assumed same as static bed
Temperature	10 to 35°C
Media	Size: 0.5 to 1.0 Density: 1.3 to 3 g/mL (low-density media preferred)
Bed Expansion	5 to 20 percent (5-10 preferred)
Flow Distribution	Uniform
Excess Solids Control	Provisions made available
Recycle	Variable, flow placed to maintain consistent bed expansion

Many laboratory studies have shown the fluid bed SMAR can meet secondary effluent standards when processing dilute wastes at nominal temperatures, and because the SMAR is inherently more energy-efficient than conventional secondary treatment, a preliminary SMAR design was developed for evaluating the relative merits of this process in greater detail. For comparative purposes, the preliminary system to be evaluated was designed to produce an effluent that meets or exceeds secondary treatment standards and to provide a well stabilized, dewatered sludge for ultimate disposal. Thus, cost and energy consumption estimates for this approach can be evaluated in relation to conventional secondary treatment processes. This alternative secondary treatment sequence is

presented in Figure 7. This preliminary design was developed for identifying the relative merits and shortcomings, the details of which are presented in Reference 3. It should be recognized that this design is for estimating costs only, and in no way is it expected to address the many critical aspects of applying expanded-bed technology, such as uniform flow distribution.

The following summarizes the results of this evaluation in which the SMAR process was compared to activated sludge and trickling filter systems, where all systems were sized for producing effluents and dewatered sludges of comparable quality. The treatment alternatives were compared on the basis of residual solids handling, cost, and energy requirements.

A comparison of sludge processing methods for the three alternative treatment systems illustrates several advantages in sludge handling realizable with the SMAR system (Figure 8). In aerobic secondary processes, a large fraction of the organic material in the sewage is converted to cellular material. Accordingly, the quantity of secondary sludge from aerobic treatment is substantial and contains putrescible organics and pathogens which require further stabilization before ultimate disposal. By contrast, anaerobic treatment in the SMAR stabilizes most of the organics in the waste by converting them to gaseous end products. As such, this process results in much smaller sludge volumes than comparable anaerobic processes. Furthermore, sludge produced in the SMAR would be relatively stable, consisting mainly of inert material and dead cells. Because of the differences in both quality and quantity of sludge produced, the use of the SMAR for wastewater treatment could potentially result in a savings in practically every aspect of sludge handling as illustrated in Figure 8.

The SMAR system also appears to be a cost-effective waste treatment alternative. As illustrated in Figure 9 costs of wastewater treatment by SMAR are comparable to conventional treatment processes. Potential savings might accrue from the combined effects of lower capital cost, lower operation requirements and greater biogas production. However, it is premature to quantify such differences at this time.

The potential energy savings is the most significant advantage of wastewater treatment with the SMAR, as illustrated by comparing total energy balances of alternative systems (Figure 10). Treatment by SMAR is estimated to consume 20 to 48 percent less energy than comparable treatment in the

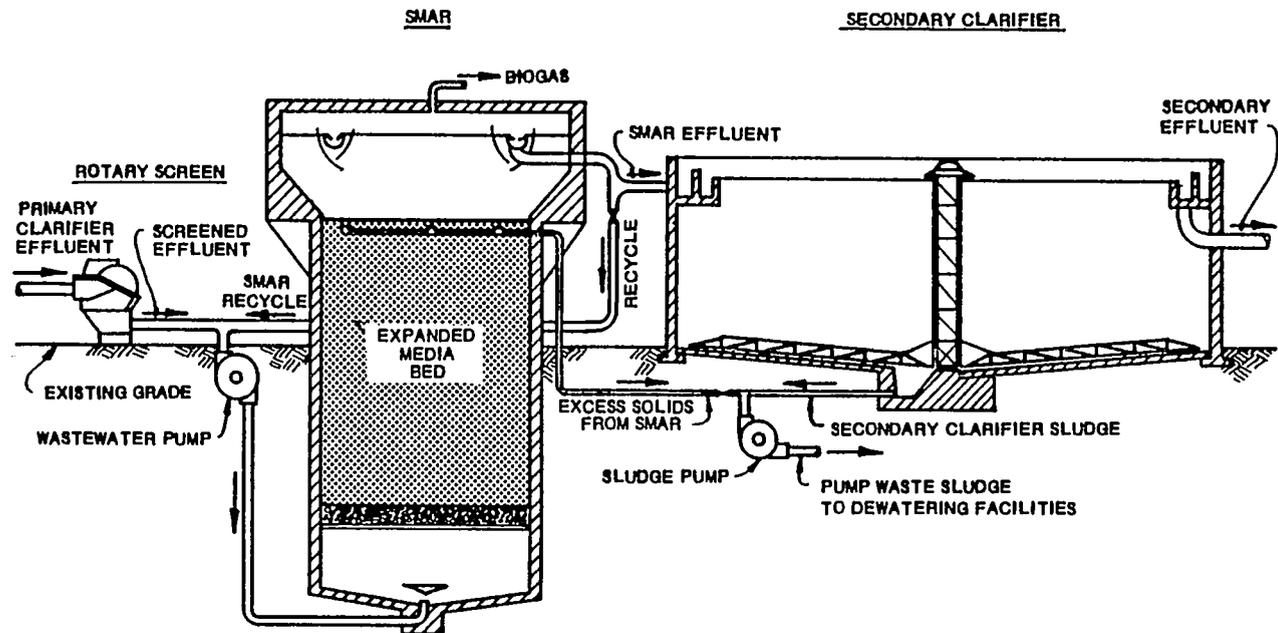


Figure 7. SMAR system for secondary treatment of municipal wastewater (3).

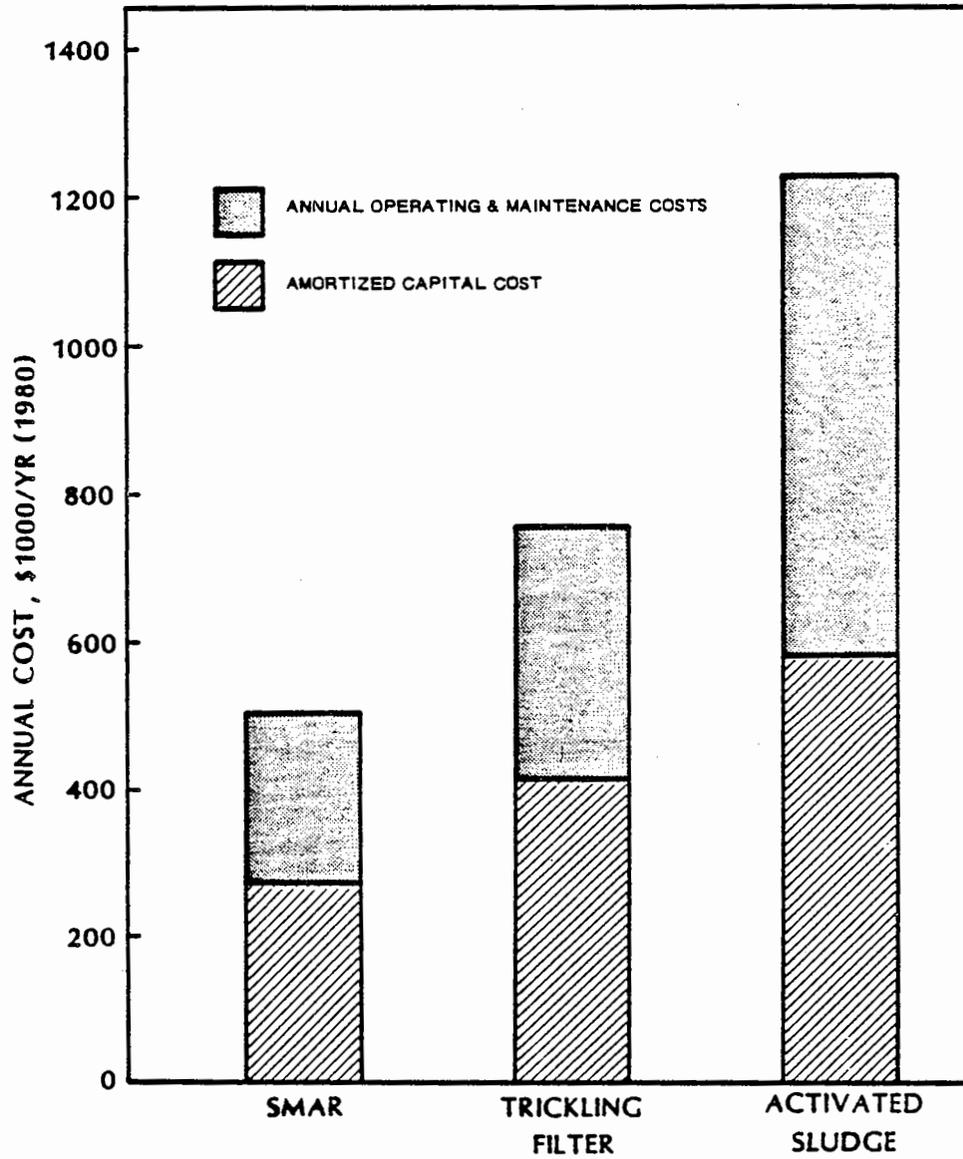


Figure 8. Residual solids handling costs for 25 mgd plant.

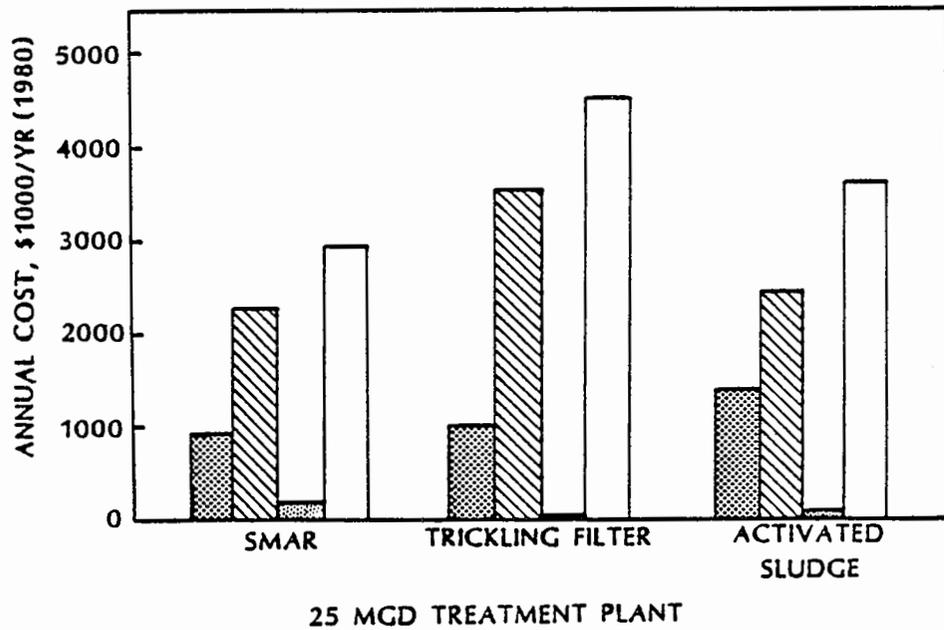
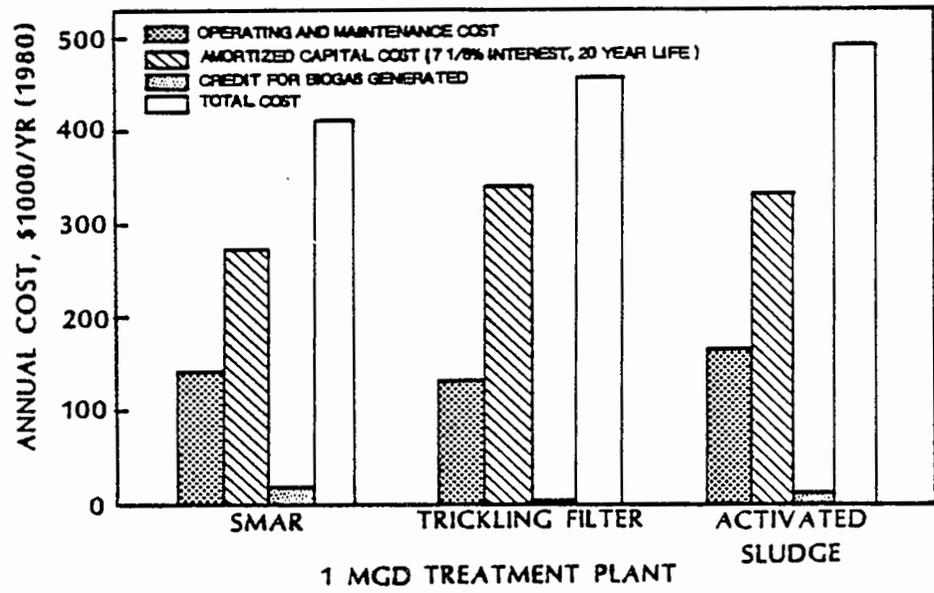


Figure 9. Summary cost comparison.

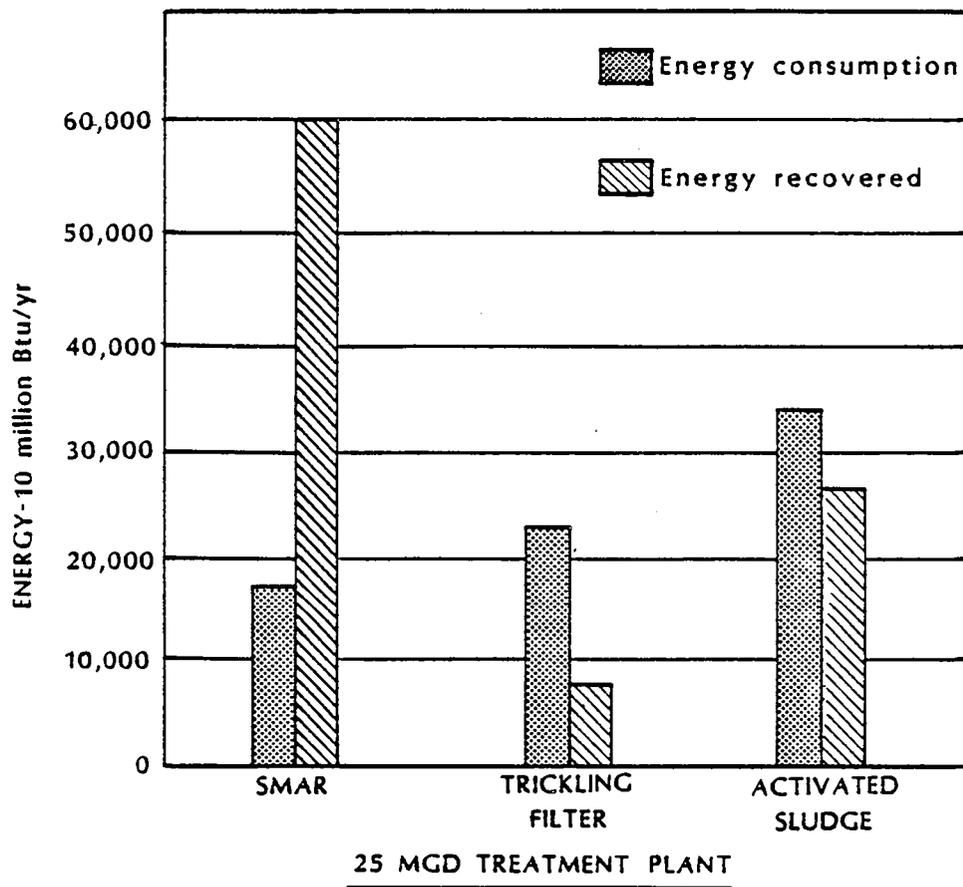


Figure 10. Comparison of energy balances for alternative secondary processes.

aerobic systems. Lower energy consumption is due to reductions in residual solids handling and elimination of energy intensive aeration. Additionally, a SMAR system could conceivably generate considerably more biogas, which can be used for on-site generation of electricity and/or any in-plant purpose requiring fuel. As shown in Figure 10 complete secondary wastewater treatment employing SMAR quite probably could be a net producer of energy through careful system design and operation.

SUMMARY

The various anaerobic processes have specific advantages and disadvantages associated with each. It is instructive to compare typical organic loading ranges of anaerobic processes with respect to conventional aerobic treatment to gain a perspective on these applications. Typical loadings of several common biological processes are compared in Table 9, where all systems are designed for 70 to 95 percent removal of biodegradable organics.

TABLE 9. COMPARISON OF TYPICAL LOADINGS TO BIOLOGICAL WASTE TREATMENT PROCESSES

Process	Organic Loading Rate, lb BOD _L /day/cu ft
Conventional Activated Sludge	0.02 - 0.075
Pure Oxygen Activated Sludge	0.10 - 0.20
Trickling Filter	0.015 - 0.30
Conventional Anaerobic Digestion	0.10 - 0.50
Standard Rate	0.01 - 0.10
High Rate	0.10 - 0.50
Anaerobic Contact Process	0.10 - 1.5
SMAR	0.10 - 1.3

When properly applied, all anaerobic processes have the potential to be net producers of energy, as opposed to alternative stabilization processes, which are essentially all net consumers of energy. Table 10 summarizes the preferred application of the anaerobic processes discussed. For all practical purposes, the energy consumption of all these processes, except the expanded SMAR, is a function of the volume of flow treated. However, for the expanded bed SMAR, energy consumption is a function of organic loadings. Anaerobic contact and conventional anaerobic processes are favorable for

TABLE 10. SUMMARY OF APPLICATION CRITERIA FOR ANAEROBIC PROCESSES

Parameter	Conventional	Anaerobic Contact	SMAR	
			Static Bed	Expanded Bed
General Application	High solids, sludges Pretreatment/sludge stabilization	High organic content, low solids Pretreatment	High organic content, moderate solids Pretreatment	Low-medium organic content, low solids Secondary treatment
Feed Solids Content	High (>3%)	Moderate (0-1%)	Moderate (0-0.5%)	Low/moderate (0-0.1%)
Organic Content	High	Moderate/high	Moderate/high	Low/moderate
Temperature	>30°C	>30°C	>20°C	>10°C
Load Fluctuations	Moderate, semi-continuous flow	Low, semi-continuous flow	Moderate/high intermittent seasonal operation	Moderate/high intermittent operation
Toxic Materials	Poor resilience	Poor resilience	Excellent resilience	Good resilience
Energy Self-Sufficient Point (Inf. BOD _L)	30,000 mg/L	10,000 mg/L	1,500 mg/L	130 mg/L
Energy Consumption	Moderate 1,000 Btu/gal	Moderate 300 Btu/gal	Low 2 Btu/gal	Low/moderate 4-100 Btu/gal

* Energy consumption a function of organic loading.

high-strength wastes. For the contact process, a soluble or low-solids content waste is preferred; whereas, high-solids wastes are best suited to conventional treatment. The static bed SMAR is excellent for pretreatment of moderate- to high-strength industrial wastes. It is particularly applicable to industries where intermittent operation is desirable, e.g., facilities that cease processing and waste production on weekends, or for seasonal operation. Furthermore, the static bed SMAR can accommodate toxic slugs quite well.

On the other hand, the expanded bed SMAR is most applicable to relatively dilute wastes where a high degree of stabilization is required, such as secondary treatment or a polishing treatment step. The expanded bed SMAR appears to be the only anaerobic process that could possibly achieve secondary effluent quality with low-strength wastes, such as municipal wastewater. Conceivably, anaerobic reactors could be sequentially staged to economically achieve a high-quality effluent with an energy-efficient system. Either the static bed SMAR, anaerobic contact, or the conventional process could be used as a roughing or pretreatment step preceding conventional aerobic treatment or the expanded bed SMAR.

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APPLICATION OF PACKED-BED UPFLOW TOWERS
IN TWO-PHASE ANAEROBIC DIGESTION

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INTRODUCTION

For many years it has been well known that biological slimes (microbial films) develop on solid-liquid interfaces in nature as dissolved and colloidal substrates are transported past the solid surfaces. Engineers have simulated this natural phenomenon in trickling filters to provide aerobic biodegradation of settled sewage. Since microbial slimes also develop in anaerobic environments in lake and river sediments and underground soils, an anaerobic counterpart of the aerobic trickling filter could also be developed to degrade organic matter to produce stabilized effluents.

The feasibility of biodegradation of organic matter in a downflow anaerobic trickling filter was indicated in 1962 by the work of Banerji (1), who showed that microbial films grown on Ottawa sand within a sealed column were able to degrade such complex organic compounds as alkylbenzene-sulfonates. Young and McCarty (2) employed laboratory-scale submerged anaerobic upflow "filters" packed with stone to gasify synthetic soluble wastes fed to the bottom of the

filters. No effluent recycle was used and the anaerobic upflow filter was assumed to behave as a plug-flow reactor.

The concept of upflow anaerobic filter is not new; indeed, these filters have been built in England since 1876 to purify sewage (3). Also, stabilization of septic tank effluent in downflow rock-filled trenches and cesspools is based on a similar principle.

PAST WORK WITH PACKED-BED UPFLOW TOWERS

Anaerobic filters are usually upflow tower reactors which can be packed with a variety of materials including stones, Raschig rings, Berl saddles, and other special packings made of plastics, ceramics, etc. Void ratios up to 90% can be obtained depending on the shape and size of the packing materials. Laboratory, pilot, and field-scale filters have been used to digest a variety of low-suspended solids industrial effluents and sewage; excellent reviews on the application of these systems were presented by Mueller and Mancini (4), Donovan, *et al.* (5), and others. The work reported in the literature indicates the following:

- Organic loading rates (LR) for these filters vary over a wide range between 0.20 and 10 kg COD/m³-day based on empty column volume.
- Hydraulic residence times (HRT) also vary widely and are between about one-half day and 80 days.
- Most filters are operated without effluent recycling.

Chian and DeWalle (6) observed that inhibition of the methane fermentation process occurred when high-strength wastes were fed to the anaerobic filter. Consequently, the filter had to be operated with diluted wastes.

Contrary to the common assumption that anaerobic filters behave as plug-flow systems, Chian and DeWalle indicated that upflow towers experiencing velocities of 1.1 m/day or more can be considered to be completely mixed. Rivera (7) showed from theoretical considerations that the anaerobic filter could exhibit flow characteristics between those of a plug-flow and complete-mix reactor depending on flow-through rates applied on the packed bed.

APPLICATION OF UPFLOW PACKED BEDS IN TWO-PHASE DIGESTION

The purpose of this research was to investigate the applicability of anaerobic upflow packed beds to serve as the methane-phase digester of an overall two-phase digestion system. In the two-phase anaerobic digestion process, the sequential acid and methane fermentation steps of the overall digestion process are optimized in separate reactors (8). Acidic end-products emanating from the first-stage acid-phase digester are ultimately converted to methane in a second-stage methane-phase digester. In this research, laboratory-scale upflow anaerobic filters were operated with acidic effluents from complete-mix acid-phase digesters fed with particulate and soluble feeds. Specifically, the objectives of the work were:

- To ascertain whether upflow anaerobic filters can be operated with feeds containing high concentrations of volatile acids (VA).
- To evaluate the effects of such operating factors as loading rate, HRT, effluent recirculation ratio, and digester temperature on the conversion of volatile acids and other products of acid-phase digestion to methane.

FILTER OPERATION WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF PARTICULATE SUBSTRATES (BIOMASS-WASTE BLEND)

Two-Phase System Feed

The first two-phase system was operated with a mixed feed containing finely ground water hyacinth, Coastal Bermuda grass, the organic fraction of municipal solid waste (MSW), and sewage sludge blended in the ratio of 1 part sewage sludge (mixture of 30 wt % primary and 70 wt % activated on a dry solids basis) to 10.5-11.1 parts hyacinth, 10.4-10.8 parts grass, and 10.6-12.6 parts MSW, all on a dry volatile solids (VS) basis. The feed had a median particle size of less than 0.5 mm. Details of growth conditions of the biomass, harvesting procedure, and processing methods to separate the organic fraction of MSW were published in a separate report (9). The moisture and VS contents of these feeds were between 85.2% and 85.6%, and 78.1% and 83.3%, respectively. Carbon, nitrogen, hydrogen, sulfur, phosphorus, calcium, magnesium, sodium, and potassium

contents of these feeds were 39.2-40.2, 1.87-1.90, 5.03-5.09, 0.35-0.37, 0.33-0.34, 1.6-1.7, 0.29-0.31, 0.86-0.99, and 0.95-1.10 wt %, respectively. The heating value of the feeds ranged between 6835 and 6859 Btu/dry lb (3797 and 3811 kg-cal/kg).

Apparatus

The two-phase system consisted of a mesophilic (35°C) 25-l (16-l culture volume) acid digester and an upflow anaerobic filter. The acid-phase digester was fed intermittently with the blend feed at the rate of 70 times/day to simulate conditions closely approaching those of continuous feeding. This digester had an HRT of 0.8 days and a loading rate of about 1.25 lb VS/ft³-day (20 kg/m³-day). The anaerobic filter was made from a 7.5-in. (19.1 cm) I.D. and 3-ft 0.75-in.-long (93.3 cm) Plexiglas column with flanged ends and was sealed at the top and bottom. It was packed with three sizes of 0.5-in. long (1.27 cm) PVC Raschig rings. The rings had I.D.'s and O.D.'s 0.71 x 0.83-in., 0.79 x 1.0-in., and 1.0 x 1.3-in. (1.8 x 2.1-cm, 2.0 x 2.6-cm, and 2.6 x 3.3 cm). The filter had a gross (empty) culture volume of 18.5 liters and a void ratio of 63%. It was fed continuously with filtered effluents from the acid-phase digester (Figure 1). The feed was stored in a magnetically mixed Belco glass reservoir operated like a Mariotte bottle to provide a constant delivery head.

The filter had three sampling ports 2.5, 12.5, and 21.5 inches (6.4, 31.8 and 54.6 cm) from the bottom of the filter. A recirculation system was installed to recycle filter effluents from the top to the bottom of the culture to dilute the incoming acidic feed, as necessary, and to enhance the transport of end products of fermentation out of the culture.

Experimental Plan

Three experimental runs, as described below, were conducted after a filter start-up and culture development and acclimation period.

- Run 1. A baseline run without effluent recirculation at 35°C.
- Run 2. Conditions were the same as in Run 1, but with continuous effluent recycle at a recirculation ratio

(ratio of effluent recirculation flow rate to daily feed flow rate) of 5.3:1.

- Run 3. Conditions were the same as in Run 2 except that the culture temperature was increased to about 38°C.

All runs were conducted at an HRT of 2.31 days based on daily feed flow rate and gross filter volume.

Anaerobic Filter Feed

Effluents from the first-stage complete-mix acid-phase digester were fed to the second-stage anaerobic filter. The mesophilic (35°C) acid-phase digester had a culture volume of 16 l; it was operated at an HRT of 0.8 days and a loading rate of about 1.25 lb VS/ft³-day (20.0kg VS/m³-day). Chemical characteristics of the vacuum-filtered acidic effluents, which were feeds for the filter, are reported in Table I for the three experimental phases designed to study the effects of effluent recycle and culture temperature.

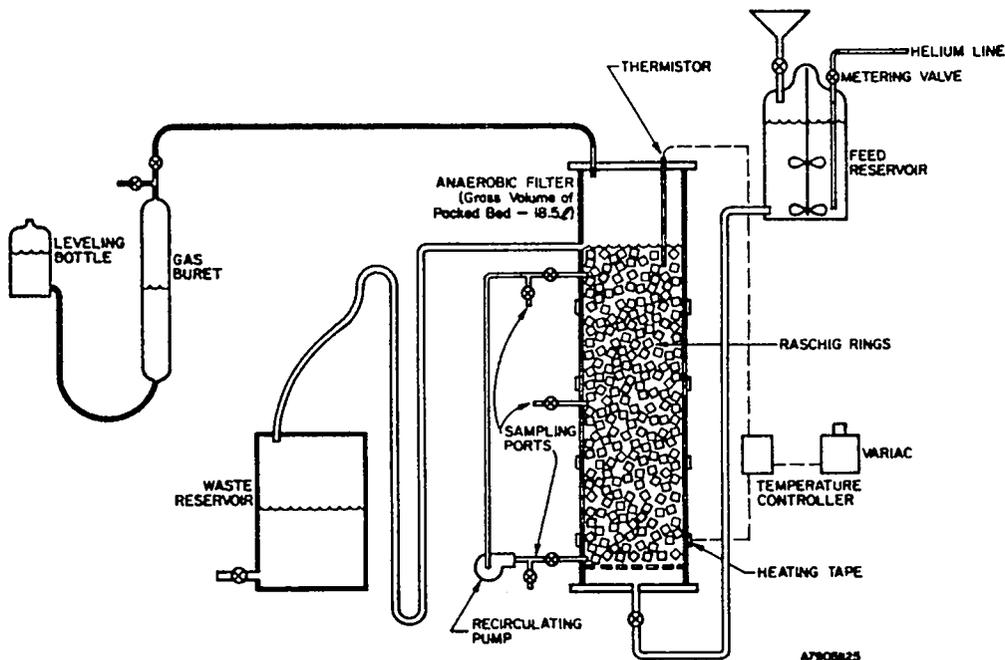


Figure 1. ANAEROBIC FILTER SERVING AS METHANE-PHASE DIGESTER OF A TWO-PHASE SYSTEM FED WITH HYACINTH-GRASS-REFUSE-SLUDGE BLEND

Table I. CHEMICAL CHARACTERISTICS OF ACID-PHASE
EFFLUENTS (Anaerobic Filter Feed)

Run	pH	NH ₃ -N (mg/ℓ)	Alkalinity (mg/ℓ CaCO ₃)		Volatile Acids (mg/ℓ)						Total	VS Conc. (g/ℓ)	
			Total	Bicarbonate	Acetic	Propionic	Butyric	Isobut.	Valeric	Isoval.			Caproic
1	5.71	38	1340	470	753	446	58	16	61	19	0	1193	3.225
2	5.41	68	1310	0	1280	598	176	24	130	21	8	2000	3.542
3	5.24	221	1270	0	1008	661	139	16	141	11	15	1747	4.658

Results and Discussion

Start-Up

The filter was started by filling it up with a mesophilic inoculum derived from the effluent of a methane-phase CSTR digester fed with acid-phase effluent, and gradually feeding it with vacuum filter supernatant from the acid-phase effluent after a period of batch operation for 3 days. As indicated by the progression of substrate conversion plotted in Figure 2, satisfactory bacterial density and methanogenic activity were established within about 26 days after filter inoculation. The methane content of the produced gases ranged between 70 and 72 mol % toward the end of the start-up phase.

The experimental runs were started after about a month of culture development and acclimation. The filter operating conditions and performance data for the three experimental runs are reported in Tables II and III.

Baseline Run

A baseline run was conducted at a standard mesophilic temperature of 35°C without any recirculation of the effluent. The data showed that the various volatile acids were converted at very high efficiencies (Table IV) even when the volatile acids concentration in the feed was about 1200 mg/l and effluent was not recirculated. As indicated by the methane concentration and yield, inhibition of methanogenesis was not experienced in this fixed-film process at this volatile acids concentration. The efficiency of gasification of non-VA volatile solids was only about 20% on a mass basis under these conditions.

Effect of Effluent Recirculation

In a methane-phase filter receiving high volatile acids-content feed, recirculation of the effluent is advisable to dilute the influent acids to a level that is not inhibitory to the methane formers. The recirculation ratio, R, can be estimated from the following equation:

$$R = \frac{C_f - C_i}{C_i - C_e} \quad (1)$$

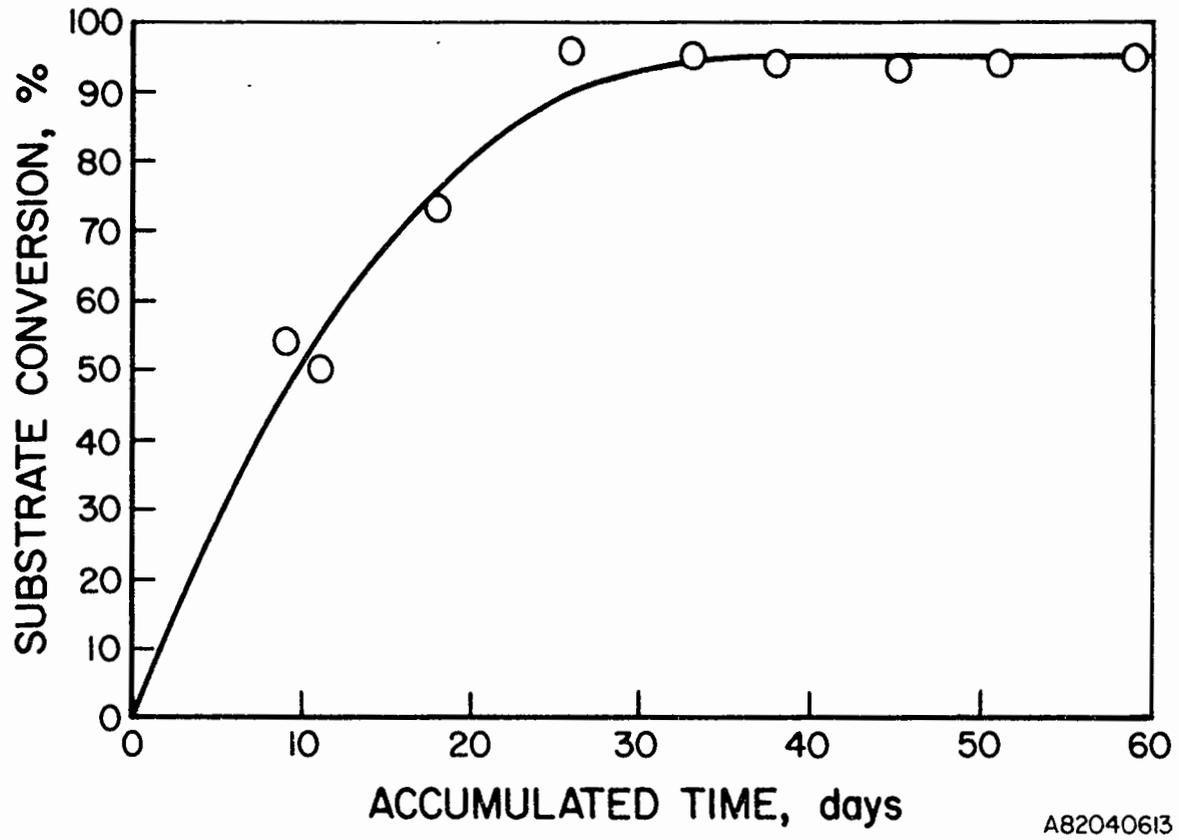


Figure II. ANAEROBIC FILTER START-UP WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF HYACINTH-GRASS-REFUSE-SLUDGE BLEND

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Table II. OPERATING CONDITIONS FOR ANAEROBIC FILTER FED WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF PARTICULATE SUBSTRATE

Run	Purpose	Temperature		Effluent Recirculation Ratio ^a	COD Loading ^b (kg/m ³ -day)		VS Loading ^b (kg/m ³ -day)		HRT ^b (days)
		Average (°C)	Range (°C)		Total	Acids	Total	Acids	
1	Baseline	35	35-35	0	1.88	1.11	1.39	0.58	2.3
2	Recirculation Effect	33	29-35	5.3	2.06	1.30	1.54	0.96	2.3
3	Temperature Effect	38	35-40	5.3	2.75	1.18	2.02	0.87	2.3

^a Recirculation ratio is the quotient of the effluent flow rate and daily fresh feed flow rate.

^b Loadings and HRT's were calculated on the basis of empty column volume of 18.5 liters.

Table III. PERFORMANCE OF ANAEROBIC FILTER FED WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF PARTICULATE SUBSTRATE

Run	pH ^a	Effluent Quality											
		NH ₃ -N (mg/l)	Alkalinity		Conductivity (μmho/cm)	Volatile Acids							Total as Acetic
			Total	Bicarb. (mg/l CaCO ₃)		Acetic	Propionic	Butyric	Isobutyric	Valeric	Isovaleric	Caproic	
1	6.88	52	1220	1180	3150	38	17	0	1	0	0	0	54
2	6.86	27	1760	1650	3050	90	41	4	2	4	4	0	131
3	6.74	310	2230	2070	6640	60	53	1	1	3	2	0	107

Table III, Cont.

Run	Gas Quality					
	Gas Production Rate ^b (vol/vol-day)	Gas Composition			Methane Yield ^c	
		CH ₄	CO ₂	N ₂	(m ³ /kg VS)	(m ³ /kg COD added)
1	0.591 (16)	73.0	27.0	0	0.31	0.23 (62%)
2	0.891 (10)	72.3	27.7	0	0.42	0.31 (85%)
3	0.721 (14)	71.4	28.6	0	0.25	0.19 (51%)

^aNo neutralizer addition was necessary to adjust digester pH.

^bGas production rate was based on empty filter volume; figures in parentheses are coefficient of variation; gas volume was at 60°F and 30-in. Hg pressure.

^cMethane yields are at standard temperature of 60°F and standard pressure of 30-in. Hg. Figures in parenthesis are percent of theoretical methane yield of 0.37 m³/kg COD.

where —

C_f = concentration of volatile acids in the system feed before dilution,

C_i = concentration of volatile acids in the influent stream to the bottom of the filter after dilution, and

C_e = concentration of volatile acids in the effluent stream.

In Run 2 the influent volatile acids concentration was 2000 mg/l. For this run, an effluent recirculation ratio of about 5 was used to ensure that the acid concentrations at the bottom of the filter did not exceed 500 mg/l compared to a concentration of about 1200 mg/l in Run 1 (Table IV). A comparison of performance data from these two runs shows that effluent recirculation had the effect of increasing the gasification efficiency of non-VA organic matter, gas production rate, and methane yield. The methane yield for Run 2 was 85% of the theoretical, the highest of these runs. Volatile acids conversion efficiencies for Run 2 were about the same as for Run 1.

Effect of Temperature

A comparison of the data in Table IV shows that filter operation at a mean temperature of 38°C did not affect volatile acids conversion, but decreased the efficiency of conversion of non-VA organics to gas. The methane yield was lowest among those runs and only about 50% of the theoretical.

Gas Composition

From stoichiometric considerations it can be shown that methanogenic conversion of a C_2 fatty acid (acetic acid) yields a gas mixture containing 50 mol % methane and 50 mol % CO_2 . Similar calculations show that higher fatty acids will yield gases richer in methane as indicated in Table V. Consequently, a mixture of C_2 and higher acids fed to a methane digester is expected to exhibit a methane concentration between 50 and 95 mol %, depending on the relative proportions of the various volatile acids. As an example, the feed for Run 1 contained 12.22 m moles/l of acetic, 6.03 m moles/l of propionic, 0.84 m moles/l of butyric and isobutyric, and 0.78 m moles/l of valeric and isovaleric acids; in theory,

Table IV. EFFECT OF RECIRCULATION AND CULTURE TEMPERATURE ON PERFORMANCE OF ANAEROBIC FILTER FED WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF PARTICULATE SUBSTRATES

Run	Influent Volatile Acid Concentration		Acid Conversion Efficiency (%)								VS Conversion to Gas (%)	Gas Production Rate (vol/vol-day)	Methane Yield	
	Before Dilution (mg/t as acetic)	After Dilution *	Acetic	Propionic	Butyric	Isobutyric	Valeric	Isovaleric	Caproic	Total			(m ³ /kg COD added)	(% of theoretical)
1	1193	1193	95	96	100	94	100	100	--	96	20	0.59	0.23	62
2	2000	429	93	93	98	92	97	81	100	94	51	0.89	0.31	85
3	1747	367	94	92	99	94	98	82	100	94	27	0.72	0.19	51

* Diluted volatile acid concentration (C_d) was calculated from the formula: $C_d = C_f (1 + Rx)/(1 + R)$, where C_f is the concentration of undiluted acid, R is the selected recirculation ratio, and x is the ratio of C_e (effluent volatile acids) and C_f .

Table VI. PREDICTED AND OBSERVED GAS COMPOSITION AND VS CONVERSION TO GAS

Run	Predicted Gas Productions From Volatile Acids Conversion								Mass of Gas in Gas Phase (g/day)	Observed Gas Production			Gas From VS (g/day)	VS Conversion to Gas (%)
	Uncorrected Total *				Corrected Gas Phase **					Total Gas (g/day)	CH ₄ (mol %)	CO ₂ (mol %)		
	CH ₄		CO ₂		CH ₄		CO ₂							
	(l/day)	(mol %)	(l/day)	(mol %)	(l/day)	(mol %)	(l/day)	(mol %)						
1	5.44	60.2	3.60	39.8	5.44	70.8	2.24	29.2	7.87	73.0	27.0	10.90	3.03	20
2	8.97	60.4	5.87	39.6	8.97	76.5	2.75	23.5	11.20	72.3	27.7	16.57	5.37	51
3	8.29	61.6	5.17	38.4	8.29	86.8	1.26	13.2	7.80	71.4	28.6	13.55	5.75	27

* Predicted gas productions (at 60°F and 30-in. Hg) were stoichiometric quantities calculated from observed conversions of various volatile acids.

** Corrected numbers are calculated assuming that methane is insoluble and that a part of produced CO₂ remains in the liquid phase in the form of bicarbonate alkalinity.

Table V. THEORETICAL GAS COMPOSITION
FROM DIGESTION OF FATTY ACIDS

Substrate	Gas Composition (mol %)	
	CH ₄	CO ₂
C ₂ Acid	50	50
C ₃ Acid	67.7	32.3
C ₄ Acid	75.0	25.0
C ₅ Acid	80.0	20.0
C ₆ Acid	83.3	16.7
C ₁₀ Acid	90.0	10.0
C ₂₀ Acid	95.0	5.0

conversion of these acids should yield a gas mixture containing about 60 mol % of methane and 40 mol % CO₂ (Table VI). Furthermore, it should be noted that a methane digester acts as a sink for the produced CO₂ because the pH conditions are such that there is an increase in bicarbonate alkalinity during methane-phase digestion of the acidic feed. Correcting for CO₂ absorption and production of bicarbonate alkalinity, predicted methane concentration in the gas phase is about 71 mol % for Run 1 (Table VI) compared with the observed concentration of 73 mol % (Table III). The methane content of methane-digester gases is thus expected to be higher than those of conventional digesters as observed during other two-phase digestion research (9-10).

Volatile Acids Profile

Data included in Table VII indicate that while there was some production and accumulation of higher acids in the bottom 10% of the anaerobic filter, conversion of the various volatile acids was essentially completed within the bottom half of the column.

Table VII. VOLATILE ACID PROFILE FOR ANAEROBIC FILTER FED WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF PARTICULATE SUBSTRATE

	Filter Height (%)	Acetic	Propionic	Butyric	Isobutyric	Valeric	Isovaleric	Caproic	Total as acetic
		(mg/l)							
Effluent, C_e	100.0	18	8	1	0	1	2	0	27
Top Port	84.1	4	1	0	0	1	0	0	5
Middle Port	48.9	4	1	0	0	0	0	0	5
Bottom Port	9.8	87	125	5	6	38	8	0	223
Filter Inlet, C_i^*	0	121	68	14	3	21	2	3	204
System Influent, C_f	--	667	387	86	16	130	5	22	1141

* Calculated from the formula, $C_i = (C_f + RC_e)/(1 + R)$ for $R = 5.3$

FILTER OPERATION WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF SOLUBLE SUBSTRATE (Soft-Drink Bottling Waste)

Two-Phase System Feed

The two-phase system was operated with a high-COD (45,000 mg/ℓ) low-suspended-solids (~76 mg/ℓ TSS, 48 mg/ℓ VSS) highly acidic waste (pH 2.5) collected from a soft-drink bottling plant. More than 99% of the total residue was organic matter. Carbon, hydrogen, nitrogen, sulfur, potassium, calcium and magnesium contents of the waste were 51.6, 5.36, 0.50, 0.07, 0.03, 0.10 and 0.004 wt % of total residue, respectively. The system feed was supplemented with NH_4Cl solution to overcome the suspected nitrogen deficiency.

Apparatus

The mesophilic (35°C) two-phase system fed with the high-strength soluble waste consisted of a complete-mix acid-phase digester (2.5 ℓ culture volume) and a 4-in. (10.2 cm) I.D. x 31-in. (78.7 cm) high upflow anaerobic filter packed with 0.83-in. (2.1 cm) long x 0.83-in. (2.1 cm) O.D. x 0.60-in. (1.5 cm) O.D. plastic Raschig rings. The Raschig rings had a bulk density of about 22 lb/ft³ (352 kg/m³) and a surface area-to-volume ratio of about 66 ft²/ft³ (218 m²/m³). The acid-phase digester exhibited a gas production rate of 1.034 vol/day per culture volume, a gas yield of 0.051 m³/kg COD added (0.82 SCF/lb COD added) under standard conditions of 60°F [15.6°C] and 30-in. [762 mm] Hg pressure), a methane content of 0.9 mol %, and a hydrogen content of 21.2 mol % when operated at an HRT of 2.2 days and a loading rate of 1.28 lb COD/ft³-day (20.5 kg COD/m³-day). The anaerobic filter had a gross culture volume of 6 ℓ (void volume 4.7 ℓ), and was fed continuously with acid digester effluents. The filter had sampling ports at 2.5, 10.5, 12.5, 20.5 and 22.5 inches (6.4, 26.7, 31.8, 52.1 and 57.2 cm) from the bottom of the filter. Filter effluent was continuously recirculated to the inlet end at a recirculation ratio of 78:1.

Anaerobic Filter Feed

Chemical characteristics of the acid digester effluent, which served as the feed for the anaerobic filter, are indicated in Table VIII. The filter feed had a high

Table VIII. PERFORMANCE OF ANAEROBIC FILTER OPERATED AT AN HRT OF 5.2 DAYS, A RECIRCULATION RATIO OF 78:1, AND A LOADING RATE OF 0.38 lb VS/ft³-day (6.1 kg COD/m³-day) WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF SOLUBLE SUBSTRATE

	Acid-Phase Effluent (Filter Feed)	Filter Effluent	Efficiency (%)
pH	4.7	7.5	--
<u>Volatiles Acids (mg/l)</u>			
Acetic	5177	219	95.8
Propionic	349	270	22.6
Butyric	3494	5	99.9
Isobutyric	6	7	--
Valeric	39	2	94.9
Isovaleric	10	5	50.0
Caproic	0	0	--
Total (as acetic)	7868	450	94.3
Ethanol (mg/l)	3540	4	99.0
NH ₃ -N (mg/l)	980	1570	00
Total COD (mg/l)	31,800	1900	94.0
Filtrate COD (mg/l)	18,380	455	97.5
<u>Volatiles Acids</u>			
COD, Calculated (mg/l)	13,150	455	97.5
<u>Non-Acid Soluble</u>			
COD (by diff.)(mg/l)	5230	0	100
Particulate COD (mg/l)	13,420	1445	89.1

concentration of volatile acids and a low pH. However, the recirculation ratio was adjusted appropriately to a high value ($R = 78:1$) to lower the acids concentration in the filter feed to non-inhibitory levels. The undiluted filter feed contained 13,420 mg/l of particulate COD, 13,150 mg/l of volatile-acids COD, and 5230 mg/l of non-volatile-acids COD due to soluble organics other than fatty acids.

Results and Discussion

Performance data in Table VIII show that the anaerobic filter could be operated at a high loading rate to convert low-pH and high volatile acids-content feeds to methane when the filter effluent was recirculated at an appropriately high volumetric rate to ensure that the organisms were not exposed to unduly large acid concentrations. Volatile acids, alcohol, and COD reductions of 94% or higher were obtained, except with propionic and isovaleric acids. Soluble organics (other than volatile acids) fed to the filter were converted to acids and finally to gases in high efficiencies. The particulate and colloidal matters, which had to be transformed to acids and then to gases within the packed bed, were 89% converted.

The anaerobic filter exhibited a methane yield of 6.57 SCF/lb COD added ($0.41 \text{ std m}^3/\text{kg COD added}$), a methane content of 68 mol %, a CO_2 content of 32 mol %, and a high gas production rate of about 3.7 volumes/culture volume-day.

The volatile acids profile (Table IX) along the filter height indicated that there were significant activities of non-methanogenic organisms within the bottom 75% of the filter; these organisms were involved in conversion of particulates to higher fatty acids and of higher fatty acids to acetate, the substrate for methane bacteria. There was evidence of good methanogenic activity within the top half of the column and particularly within the top 25% of the filter. Also, it appeared that higher acidogenic and methanogenic activities were localized near the dispersion plates installed at 37.5% and 72.5% of the column height.

CONCLUSIONS

The following conclusions can be drawn based on information collected during this research:

Table IX. VOLATILE ACIDS PROFILE FOR ANAEROBIC FILTER FED WITH EFFLUENTS FROM ACID-PHASE DIGESTION OF SOFT DRINK BOTTLING WASTE

	Filter Height (%)	Acetic	Propionic	Butyric	Isobutyric	Valeric	Isovaleric	Caproic	Total as Acetic
		(mg/l)							
Effluent, C_e	100.0	442	121	18	18	10	8	0	575
Port 5	72.4	435	158	11	15	10	8	0	591
Port 4	66.1	451	167	360	36	15	8	0	871
Port 3	40.3	513	170	425	59	21	5	0	998
Port 2	33.9	679	194	104	23	20	8	0	940
Port 1	8.1	296	133	8	5	0	0	0	413
Filter Influent, C_i^*	0	474	122	110	18	10	8	0	671
System Influent, C_f	--	3009	219	7262	8	18	12	0	8160

* Calculated from the formula $C_i = (C_f + RC_e)/(1 + R)$ for $R = 78$.

Methane-Phase Filter Receiving Effluents From Acid-Phase Digestion of Particulate Feeds

- Upflow packed-bed anaerobic filters can be used as methane digesters to convert influent volatile acid concentrations up to 1200 mg/l to methane without inhibition of the methane bacteria. Conversion efficiencies of 94% to 100% and 20% were obtained for volatile acids and organic solids without any effluent recirculation at 35°C at an HRT of 2.3 days and a loading rate of about 0.12 lb COD/ft³-day (1.9 kg COD/m³-day). Methane yield under these conditions was 3.65 SCF/lb COD added (0.23 m³/kg COD added, about 62% of the theoretical yield), and the gas production rate (GPR) was 0.6 vol/culture vol-day.
- Effluent recirculation was desirable when the filter feed contained volatile acids concentrations higher than 1200 mg/l. Recirculation had the effect of substantially increasing the gas production rate, methane yield, and gasification efficiency of non-VA organics. Methane yield with recirculation was 4.97 SCF/lb COD added (0.31 m³/kg COD added, 85% of the theoretical yield). Efficiencies of acids conversion with the 2000 mg/l acids-content feed and effluent recirculation were about the same as those with the 1200 mg/l acids-content feed and no recirculation.
- Increase of filter temperature by 5°C did not affect acids conversion but decreased the efficiency of conversion of non-VA organics.
- Gases from methane-phase filters are expected to exhibit methane concentrations significantly higher than those from conventional digesters because they are fed with mixtures of low oxygen-content feeds, and because there is an increase in bicarbonate alkalinity during the methane fermentation process.
- For low-COD loading rates and feeds containing particulate substrates of low biodegradability, gasification of acids and other organics is essentially completed within the bottom half of the filter.

Methane-Phase Filter Receiving Effluents From Acid-Phase Digestion of Soluble Feed

- An anaerobic filter could be operated as a methane-phase digester at a high loading rate of 0.38 lb COD/ft³-day (6.1 kg COD/m³-day) with a high volatile acid-content (~8000 mg/l as acetic) feed when the effluents were re-circulated to appropriately dilute the incoming acids. A high GPR of 3.7 and methane production in theoretical yields were obtained at an HRT of 5.2 days.
- Conversions of 90% to 100% of the soluble and particulate organic matters were obtained.
- For this case where a large part of the filter feed was non-VA soluble and particulate organics of high biodegradability, there was evidence of significant activities of nonmethanogenic organisms within the bottom 75% of the filter.
- Methanogenic activities were dominant within the top half of the filter.

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PART XIII: INDUSTRIAL WASTEWATER TREATMENT

PERFORMANCE CHARACTERISTICS OF ANAEROBIC
DOWNFLOW STATIONARY FIXED FILM REACTORS

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ABSTRACT

Stationary fixed film reactors operated to ensure a net downflow of substrate have several characteristics different from other retained biomass reactors.

1. The active biomass attaches itself to stationary surfaces and hence is difficult to wash out.
2. Performance is related to the surface-to-volume of the film support as well as to the composition of the support. Methane production rates of up to $8 \text{ m}^3/\text{m}^3/\text{day}$ at loading rates of up to $30 \text{ kg COD}/\text{m}^3/\text{day}$, are possible.
3. Severe hydraulic and organic overloadings can be tolerated with operation back to normal 24 hours following cessation of mistreatment.
4. Reactors can operate with dilute and concentrated wastes ($4000\text{--}130,000 \text{ mg COD}/\text{L}$) and can change readily over from one waste to another.
5. Intermittent loading at high loading rates are possible.
6. Methane production rates and loading rates decreased linearly with temperature (35 to 10°C); at 10°C they were about 20% of those at 35°C .

NRCC No. 20071

INTRODUCTION

Interest in anaerobic fixed film reactors has increased rapidly over the last decade with the realisation that cost-effective anaerobic treatment is generally not possible without some form of retention of micro-organisms (14). Advanced anaerobic reactors which are based on retention, although they are not all or exclusively fixed film reactors, include: the contact process reactor, the upflow sludge bed reactor, the upflow packed bed reactor (filter), the fluidised or expanded bed reactor and the downflow stationary fixed film (DSFF) reactor. These reactors are finding commercial application because the retention of the micro-organisms eliminates or reduces some of the problems associated with older systems: low rates of performance, instability, inability to withstand hydraulic and organic shockloads and extreme sensitivity to toxic materials. Fixed film reactors have advantages over other retained-biomass reactors in that the micro-organisms are attached to inert material and therefore easier retained.

The DSFF reactor was developed at the National Research Council of Canada to further address two problem areas of many advanced reactors. These reactors tend to form local high concentrations of volatile acids with concentrated wastes at the high densities of micro-organisms used. As a result the waste needs to be diluted (with water or effluent) or mechanical agitation or pumping need to be used to avoid local high concentrations of substrate. In the second place, these reactors remove effluent at the top of the reactor resulting in an accumulation of suspended solids in the reactor. This eventually causes plugging and loss of performance. Solutions to these problem areas were obtained by choosing a downflow configuration. This means adding the fresh substrate to the reactor at the spot where mixing caused by the gas production is most intense. As a result, high concentrations of substrate are immediately dispersed and there is little need for an elaborate substrate distribution system. The downflow mode also requires removal of effluent and suspended solids from the bottom of the reactor, thereby preventing plugging. The downflow mode of operation required a stationary film support to maintain the film of micro-organisms in the reactor. To prevent settling of suspended solids on parts of the film support surface, the stationary film support had to be organised in vertical channels.

This paper describes and summarizes results obtained in the last four years with DSFF reactors. Most of the work was done at the laboratories of the National Research Council of Canada (4-7,9,10,12-24). Larger scale work was done at the research laboratory of Canadian Cannery Limited (11) and at the Wastewater Technology Center of Environment Canada (2). Results include those from studies on the effects of support materials, type and concentration of wastes, size of reactor, temperature, method of waste addition (continuous versus slug loading) and shock or overloading. Also observations on start-up and nutrient requirements are discussed.

REACTOR DESIGN, OPERATION AND STUDIES

The method of operation of DSFF reactors can be most readily understood from Figure 1 showing a multiple channel DSFF reactor. Waste was pumped in at the top, together with recycled effluent when desired, and effluent was withdrawn from the bottom. This method of operation ensured that suspended solids, including growth, did not accumulate in the reactor. Bacteria from the inoculum attached themselves to the inside channel walls of the film support material and grew there. The loading rate and ultimate film thickness depended on channel surface characteristics and composition and on waste composition. Waste organics and nutrients diffused into the film and were converted into methane and carbon dioxide. The release and escape of gas from inside the film assisted mass transfer into the film, but also caused vigorous liquid movement. The gas lift action caused liquid to go down some channels and up others depending on differences in activity between channels. Overall performance depended on the total amount of film biomass present and hence on the surface-to-volume ratio of the support material. Much of the laboratory results were obtained with single and 4-channel reactors.

Film support materials tested included glass, foamed and solid polyvinyl chloride, two types of fired clay (ceramics), and needle punched polyester (13,18). These materials were compared in single and 4-channel reactors (0.6 to 1.2 L) using bean blanching waste as substrate at 35°C.

Comparisons included rate of start-up as well as maximum steady state performance.

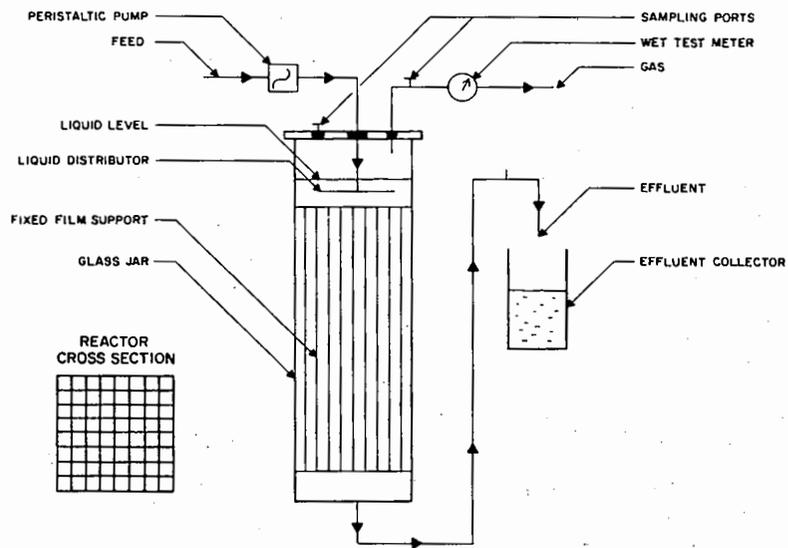


Figure 1: Set-up of multi-channel downflow stationary fixed film digester

TABLE I

Composition of Wastes Treated in Downflow Stationary Fixed Film Reactors

Waste	Waste strength ^a (total COD) g/L	Suspended COD, % of total	Sodium, g/L	Kjeldahl nitrogen g/L	Total phosphate gP/L	Ratios	
						COD/N	COD/P
Barley Stillage Waste	53	25	-	1.1	-	48	-
Bean Blanching Waste	10 (4-40)	10-30	-	0.4	0.1 ^d	25	100
Chemical Industry Waste	14	0	-	2.5	0.3	5.6	47
Citric Acid Waste	3.6	-	-	0.04	0.025	90	140
Dairy Waste	4	< 15	0.58	0.2	0.06	20	67
Heat Treated Sewage Digester Sludge Liquor (HTL)	10.5	< 10	-	0.8	0.1	13	100
Pear Peeling Waste	130 (110-140)	35-50	0.4	2.3 ^c	0.47 ^d	55	275
Piggery Waste	39 (27-51)	60-70	-	2.9	0.8	13	49
Rum Stillage Waste	60 (50-70)	< 10	0.7	1.1 ^c	0.21 ^d	55	285
Skim Milk Waste	4	0	0.23 ^b	0.2	0.04	20	100
Sugar Waste	10	0	-	0.29	0.09	35	108
Synthetic Sewage Sludge	55	85	-	2.7	0.32	20	170
Tomato Peeling Waste	15 (15-30)	< 20	2.2	0.28	0.2	54	75
Whey	66	< 5	0.33	0.65	0.65	100	100

a Average or most common concentration used; range used in brackets.

b Sodium hydroxide added to increase alkalinity.

The types of wastes studied varied widely in strength, suspended solids content and types of material present (Table I). Soluble wastes included a synthetic sugar waste, food processing wastes, dairy wastes and wastes produced by chemical industries. Wastes containing substantial amounts of suspended solids included a synthetic sewage sludge, food processing wastes and piggery waste. Nitrogen and phosphate (as NH_4 - bicarbonate and K and Na-phosphate) were added where these nutrients might be in short supply.

Process variables studied included temperature (35 to 10°C), recirculation (mostly 0 and 4 times the feed rate), size of the reactor (0.65 to 650 L), intermittent (as opposed to continuous) feeding, hydraulic overloading and starvation (for up to 5 months). Most of these variables were studied with bean blanching waste and a clay support reactor. Some tests were also made with liquor from heat treated sewage sludge. Except for temperature effect studies, all tests were made at 35°C.

Fermenters were inoculated with sewage digester sludge or liquid from an active laboratory digester. Substrate was added to maintain effluent volatile acids levels between 200-600 mg/L (16, 18, 24), except for piggery waste when acids levels were usually around 1,000 mg/L.

Mesurements and analyses made during start-up, and steady-state conditions were mostly those standard for anaerobic digestion studies: feed rate and composition, gas production rate and composition, Chemical Oxygen Demand (COD), total, volatile and suspended solids contents, methane and carbon dioxide contents, Kjeldahl and ammonia nitrogen contents, total and soluble phosphate contents, volatile acids and alkalinity (16, 17). In some instances, methanogenic activity of fermenter liquids were also determined (3, 10, 12, 22).

REACTOR PERFORMANCE AND CHARACTERISTICS

Start-up Observations

Start-up phenomena of anaerobic reactors can be attributed to three main factors:

1. Quality of the inoculum in terms of quantity of micro-organisms adaptable to the waste.
2. Rate of adaptation of these micro-organisms to the waste.
3. Rate of growth of micro-organisms during and after adaptation in relation to loss of micro-organisms in the effluent.

Little is known about what determines a good inoculum and about the adaptation process.

For example, which types of bacteria require the most adaptation? Is adaptation a question of establishing qualitatively new ecological relationships or of quantitative adjustments to existing relationships? Acetic acid converting methanogens are most certainly involved because their slow growth rates limit start-up.

Experience with anaerobic digesters generally has shown that there are large differences in start-up as a result of the inoculum (1, 8, 14, 20, 25). With DSFF reactors similar observations were made. Sewage digester sludge generally required a longer time to adapt than effluent from an active digester fed food processing waste. Well balanced wastes provided a faster rate of start-up than unbalanced wastes. For example, reactors were difficult to start up with chemical industry waste (toxicity could not be demonstrated), while reactors started readily on food processing wastes or sugar waste.

Support material affected rate of start-up markedly (Figure 2). Reactors made from rigid foamed polyvinyl chloride could not be started at all, presumably because of toxicity. Glass reactors were slow to start up, presumably because bacteria had difficulty attaching themselves to the smooth inert surface (18, 19). Solid polyvinyl chloride

(PVC, used extensively in biological waste water treatment) was substantially better than glass as a film support, but not as good as the fired clay and needle punched polyester (13). Surface roughness appeared to play a major role (and in the case of clay, availability of minerals leaching out). Detailed tests using acetate-converting methanogens showed the importance of support roughness and composition for the attachment of these bacteria (9, 10).

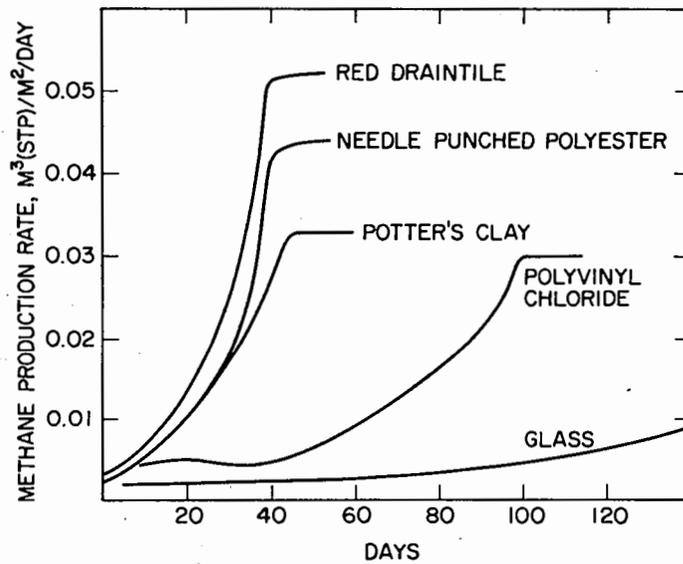


Figure 2: Actual start-up of stationary fixed film reactors (35°C; bean blanching waste, 10g COD/L; reactor volume 0.8 - 1.2 L; surface area-to-volume ratio, 100 - 150 m²/m³).

The rate of development after initial adaptation is a function of the net growth rate [total rate (depending on waste composition for a given temperature (12)) minus the fraction of the bacteria lost in the effluent]. For a model system, a mathematical relationship can be derived between growth rate, inoculum strength after adaptation, feed strength and fraction of the bacteria retained in the reactor. Typical results of such a calculation are shown in Figures 3 and 4, for reactors without and with retention of micro-organisms respectively. Results such as are presented in these figures suggest that:

1. Rates of start-up of reactors without biomass retention are markedly dependent on growth rate and initial inoculation quality (the latter is reflected by the initial loading rate).
2. The rate of start-up of reactors with biomass retention is much faster than those without retention and depends markedly on the fraction of bacteria lost in the effluent.

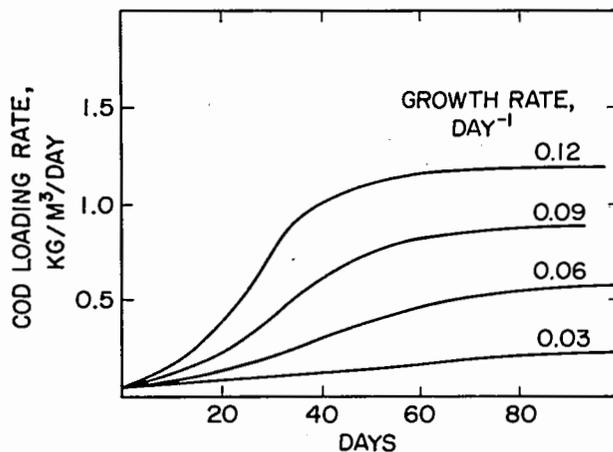


Figure 3: Calculated rates of start-up for a completely mixed anaerobic reactor without biomass retention (waste strength, 10g COD/L).

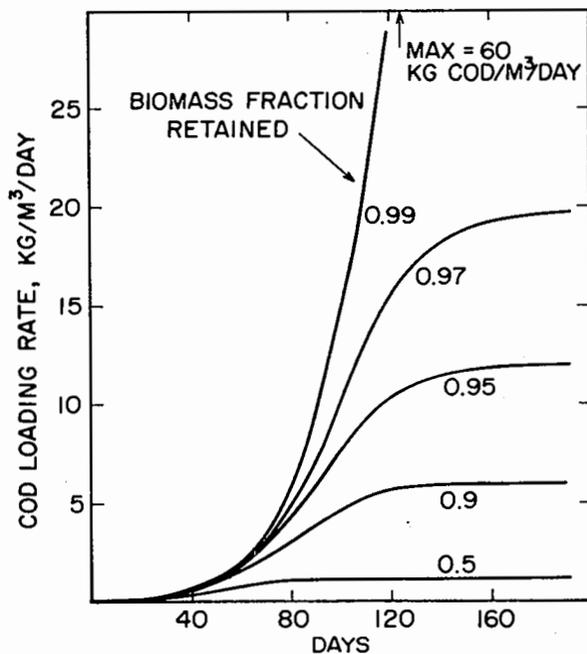


Figure 4: Calculated rates of start-up for a mixed reactor with retention of biomass (waste strength, log COD/L; growth rate, 0.06 day^{-1}).

Comparison of Figure 2 with Figure 4 shows that support material may affect both growth rate and the fraction of bacteria lost in the effluent and that in practise loading rates level off more sharply than the model suggests. The latter may indicate the effect of diffusion resistance in the film once a certain film thickness has been reached.

Waste Characteristics and Steady-State Performance

Methane Production and COD Removal

Accurate measurements of methane production during steady state operation of fixed film reactors with many wastes has shown that the volume of methane produced (at STP, i.e. 0°C, 1 atm) was equal to $0.33 \pm 0.02 \text{ m}^3$ per kg COD removed. This figure is below the theoretical value of 0.35, presumably because of methane dissolved in the effluent, accumulation of biomass in the film (in spite of "steady state" conditions) and inconsistencies in COD measurements between waste and effluent. For the purpose of this paper, therefore, methane production and COD removed (destroyed) are used interchangeably using the conversion value of $0.33 \text{ m}^3/\text{kg}$.

Waste Strength

Fixed film reactors could readily be operated at high loading rates with waste strength between 4 and 140 g COD/L (Table II, III). The maximum loading rate and the percentage of COD removed generally increased with waste strength. Both are presumably affected by the hydraulic retention time which decreases markedly with decreasing waste strength.

Suspended Solids Content

Suspended solids content affected rates of methane production to the extent these solids could be destroyed during the relatively short residence times of the substrate (Table III). Cellulose, when finely divided, appeared to be broken down quite rapidly in some wastes (23, 24). The effect of suspended solids content was particularly noticeable with pear peeling waste and piggery waste where solubilization of suspended solids was the rate limiting step rather than the rate of acid conversion to methane (5,7). The latter was most important with soluble wastes as indicated by the rapid increase in acetic acid levels when increasing loading rates (11).

Nitrogen and Phosphate Content

Maximum rates of methane production did not appear to be noticeably affected by nitrogen and phosphate contents of

TABLE II

Effect of Waste Strength on Performance of Downflow
Stationary Fixed Film Reactor at 35°C (11).
(106.5L void volume; PVC packing ("KORO-Z"); 150 m² area per m³)

Waste	Waste strength g COD/L	Loading rate, kg COD/ m ³ /day	Methane production m ³ (STP)/m ³ /day
Bean Blanching Waste	6	7.8	1.8
	12	8.3	2.2
	24	8.4	2.4
Tomato Peeling Waste	15	16.7	2.9
	30	16.9	2.8

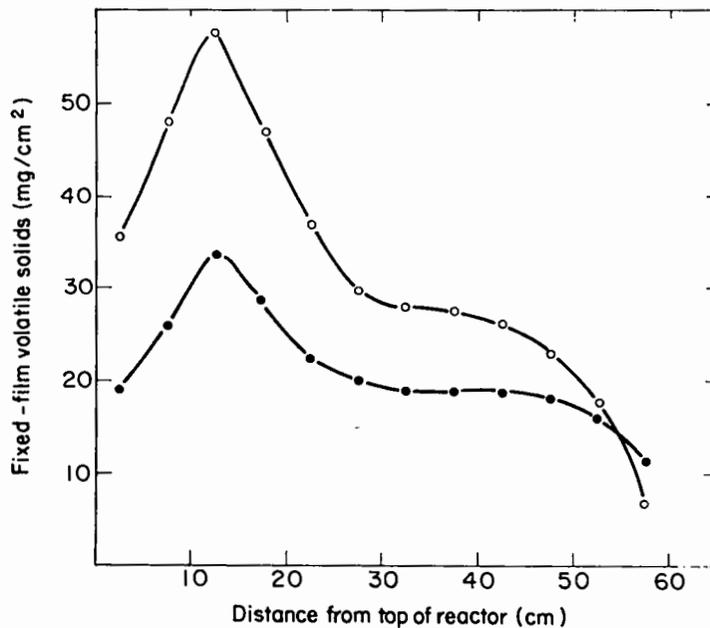


Figure 5: Distribution of volatile solids on polyester support material in DSFF reactors treating bean blanching waste (10g COD/L); ●, control; ○ nickel, cobalt and molybdenum addition.

TABLE III

Effect of Waste Strength, Suspended COD Contents and Hydraulic Retention Time on COD Removal and Rate of Methane Production in a Downflow Stationary Fixed Film Reactor^a at 35°C (21).

Waste	Waste Strength, g COD/L	Suspended COD, % of Total	Hydraulic retention time, days	Loading Rate, kg (COD)/m ³ /day	COD removal, %	Methane Production, $\frac{\text{m}^3 \text{ (STP)}}{\text{m}^3 \text{ day}}$
Bean Blanching Waste	10	10-30	1.5	6.7	84	1.9
	10	10-30	0.85	11.6	86	3.3
Piggery Waste	27-51	60-70	8.0	6.1	70	1.4
	27-57	60-70	2.7	14.5	43	2.2
	27-57	60-70	1.0	39.2	27	3.8
Pear Peeling Waste	110-140	35-50	17.5	6.4	58	1.2
Rum Stillage Waste	50-70	<10	4.5	13.3	57	2.5
Synthetic Sewage Sludge	55	85	7.4	7.4	77	1.9
	55	85	4.0	13.8	71	3.3

a 35 L void volume; made from potter's clay and consisting of vertical channels, about 2.8 X 2.8 cm inside cross section. Reactor had a surface-to-packing void volume of about 157 m²/m³. Reactors operated without recirculation, except for piggery waste.

waste substrates (compare results in Tables I to III). It is generally assumed that COD:N:P ratios of 100:5:1 are optimum. However, results presented here show that COD:N:P ratios in the ranges 100:(10-1):(5-0.1) were acceptable. Also, free ammonia nitrogen levels as high as 3500 mg/L in the effluent did not appear to affect performance.

These results show that DSFF reactors are different in nutrient requirements than suspended growth reactors. In the latter, effective growth rate and performance are related to nitrogen content (20).

Mineral Content

Limited tests on the effect of iron, nickel, cobalt and molybdenum (3, 9, 10, 22) on fixed film development showed that for some wastes addition of these minerals would improve fermenter development and performance. The effect appears to be the development of a thicker film rather than a faster rate of formation of the film (Fig. 5). To demonstrate the need for mineral addition for a given waste will require more extensive studies.

Effect of Reactor Size and Operation on Performance

Reactor Size

Only limited information is available on the effect of size on performance, because comparative studies with the same support material have not been made. A study of the data assembled in Table IV, however, indicates that all reactors (10.6-600 L void volume, 0.3-1.8 m tall, single and multiple channels) were capable of loading rates in excess of 10 kg COD/m³/day (over 3 m³ CH₄ (STP)/m³/day). Qualitative observations also indicate that in many cases multichannel, tall reactors perform as well as or better than shorter reactors (2). Full scale tests will be required to establish this.

Waste Substrate Flow Direction

The downflow mode of operation was more trouble-free than the upflow mode (11). On the other hand, upflow operation may result in a combination of a fixed film and a sludge bed reactor with higher contents of biomass and higher

TABLE IV

Effect of Reactor Size on Performance of Downflow Stationary Fixed Film Reactors^a at 35°C (2, 21)

Waste	Waste Strength, g COD/L	Reactor Size (void volume), L	Support material	Loading rate, kg(COD)/m ³ /day	COD removal, %	Methane production, m ³ (STP)/m ³ /day
Bean Blanching Waste	10-12	0.8-1.3	PVC	9.5	93	2.9
				16.9 ^b	93	5.2
		106.5	PVC	10.3 ^b	78	2.8
		0.8-1.3	Clay	10-11	87-92	2.8-3.3
				18-26 ^b	88-91	5.4-7.9
35	Clay	11.6 ^b	86	3.3		
HTL	10.5	0.8-1.3	Clay	29.2 ^b	70	6.8
		650	Clay	19	70	4.2

a Surface-to-volume ratio is 150 m²/m³ for 0.8-1.3, 35 and 110 L reactors, 70 for 650 L reactor.

b Maximum possible loading rate.

rates of methane production (14, 18, 19). Limited test with a horizontal direction of flow indicated that higher rates of performance were possible with the downflow configuration.

Recycling of Effluent

DSFF reactors were able to produce methane at high rates (over $3 \text{ m}^3/\text{m}^3/\text{day}$) with and without recycling of effluent. This was valid with dilute and concentrated wastes (4-140 g COD/L). In contrast, suspended growth systems require effluent recycling at waste strengths of about 8 g COD/L or higher (14, 21).

With waste containing large amounts of easily settleable suspended solids, such as pear peeling waste and piggery waste, a higher level of recirculation appears to be beneficial. Recirculation maintains these solids in suspension and brings them into contact with the film, thereby facilitating their breakdown (6). Tests to determine the need for recirculation in more detail are in progress.

Temperature

Performance of fixed film reactors increased linearly with temperature in the range 10 to 35°C. The effect of temperature on micro-organisms and enzymes is usually exponential in nature. The linear effect may therefore be a result of several factors.

1. Film biomass may increase with decreasing temperature (biomass yield coefficients increase with decreasing temperature (12)).
2. Diffusion resistance into the film may decrease less with temperature than microbial activity.
3. Methanogenic bacteria not in the log phase of growth have activities that increase approximately linear with temperature (12).

The linear effect of temperature on growth indicates that DSFF reactors can be used at ambient temperatures, with a relatively small penalty in performance.

Non-Steady State Performance

Change-over From One Waste to Another

DSFF reactors could readily be changed over from one waste to another or from one waste concentration to another (7, 11). These change-overs could be made with little or no loss in rate of methane production. This was also the case when changing from a low (≤ 20) to a high (4000 ppm) sodium waste and when changing from a food processing to a chemical industry or sewage treatment waste.

Hydraulic Overloading

DSFF reactors readily coped with hydraulic overloading without loss of activity (4,6). Reactors were back to normal 24-48 hours after overloading to up to 90 kg COD/m³/day at 35°C from a steady state loading of 10 kg COD/m³/day. During overloading methane production rates increased up to 15 m³ (STP)/m³/day. Similar results were obtained at 10°C (4).

Intermittent Loading

Once or twice a day (from 0.1 to 7 hours out of 24) loading increased the maximum possible daily methane production rate by close to 30% (15). During short time loading large amounts of hydrogen were produced in addition to methane (Fig. 6,7). Presumably hydrogen-converting methanogens were unable to assimilate all the hydrogen produced in such a short time. No adverse effects of continued intermittent loading on reactor performance were observed. Intermittent loading can therefore be used to reduce storage capacity for raw waste and to tailor methane production to the energy needs of the plant.

Starvation

Reactor starvation for long periods (up to several months were studied in the laboratory) affected subsequent performance for only a short time. Mostly acidogenic bacteria appeared to be affected (15) and reactors were back to high loading rates (over 10 kg COD/m³/day) within 48 hours.

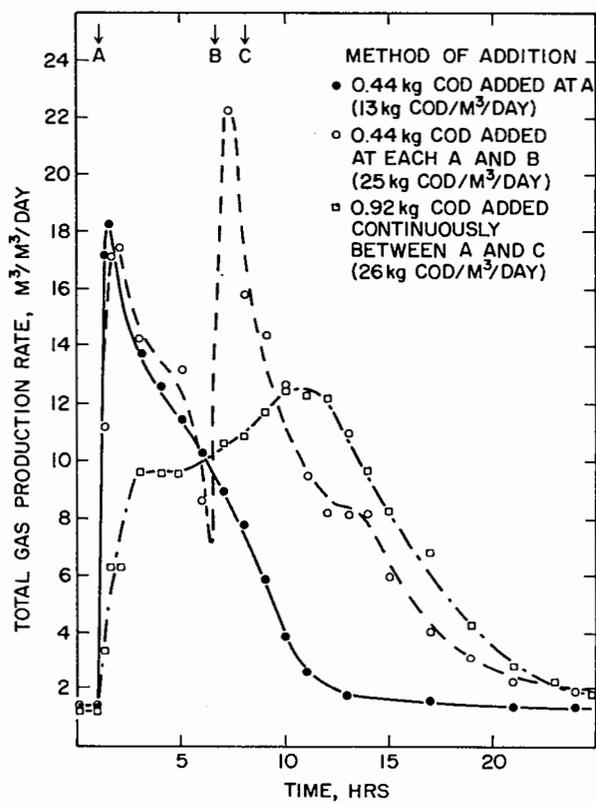


Figure 6: Effect of substrate addition method on rate of gas production (35L DSFF reactor, pear peeling waste).

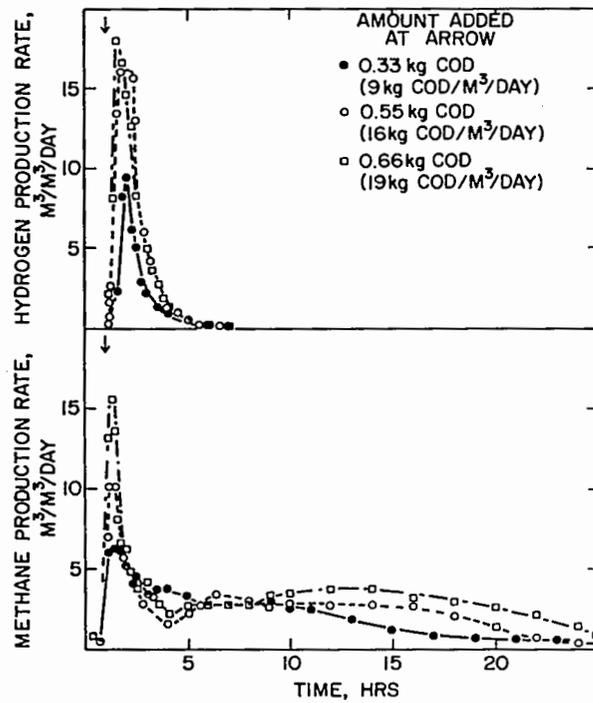


Figure 7: Effect of amount of substrate added (at arrow) on rates of methane and hydrogen production (35L DSFF reactor, pear peeling waste) during slug loading.

DISCUSSION

The most remarkable feature of the downflow stationary fixed film reactor is its ability to maintain high rates of methane production under adverse conditions. The stationary fixed film appears to provide a high degree of protection for active biomass. As a result, the DSFF reactor can withstand low temperatures, severe and repeated hydraulic overloadings, organic shockloads, sudden changes in waste composition, and starvation with little or no effect on subsequent performance. Also, high rates of methane production can be obtained while tailoring methane production to energy needs. Few, if any, retained biomass reactors can match the DSFF reactor in these respects.

Measurements of film thickness and suspended solids contents have indicated that the film is thickest near the top of the reactor (10). Acid and COD analyses of the liquid at various points in reactors have shown excellent mixing of the reactor contents. Nevertheless, mixing was not sufficient to provide uniform growth. Possibly taller reactors would improve the mixing further. Film thickness of a mature reactor generally varied from 1-4 mm and the biomass activity of the film varied from 0.8-1.5 kg COD/kg volatile suspended solids/day (0.25-0.5 m³ CH₄ (STP)/kg/day).

Studies with the DSFF reactor are continuing to obtain further information on support material and configuration, temperature, recirculation, nutrient requirements and start-up. Negotiations are also underway to obtain evaluation of DSFF reactors on a commercial scale. Many of the performance characteristics are, or may be, scale dependent, and satisfactory answers can only be obtained by a full-scale evaluation.

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TANNERY EFFLUENT: A CHALLENGE MET BY
ANAEROBIC FIXED FILM TREATMENT

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INTRODUCTION

Tannery wastewaters are extremely difficult to treat relative to most municipal wastewaters due to both the presence of very high levels of organic matter and the nature and quantity of chemical additives used in converting animal skins to leather. Proposed EPA discharge standards for the tanning industry include effluent limitations requiring the application of the best available technology economically achievable (BAT) to produce effluent discharges containing less than 40 mg/l BOD₅ and 250 mg/l COD (1). At the present time (spring 1982) few, if any, tanneries using hair pulp beaming processes for cowhides can routinely achieve these wastewater treatment goals.

The major fractions of both the organic load and flow generated at a typical tannery result from the "beaming" operations involved in preparing raw hides for tanning. These operations include washing and trimming the raw hides, removal of hair and excess flesh, and chemically preparing the hides to accept tanning agents (reliming). The unhairing step usually

requires 1-2 percent concentrations of lime and "sharpening agents" such as sodium sulfide or sodium sulfhydrate, and results in high concentrations of soluble and particulate hair protein (keratin). As a result of these processes, beamhouse wastewaters contain large quantities of lime, sulfide, dissolved and particulate hair protein and lesser quantities of salt, flesh particles, fat, blood and manure. Each of these beamhouse wastewater constituents contributes to the difficulty of meeting the proposed effluent criteria in a cost effective manner by conventional wastewater treatment methods.

Figure 1 illustrates the slow biodegradability of a pretreated beamhouse wastewater sample prior to and following treatment by an anaerobic filter. Based on these data, it is obvious that BOD_5 is a poor parameter for evaluating either beamhouse wastewaters or treatment processes due to the slow aerobic breakdown and metabolism of beamhouse wastewater constituents.

This study was conducted as part of a project designed to evaluate the feasibility of using an anaerobic process for roughing treatment of beamhouse wastewaters and the sequential use of conventional aerobic processes for final polishing treatment. Bench scale anaerobic filters were originally chosen as the reactor configuration (1979) because biomass attachment and/or entrapment would help assure long biomass retention times in the reactor and resistance to shock and toxic loads. Laboratory scale rotating biological contactor (RBC) units were used for comparing aerobic treatment of both pretreated beamhouse wastewaters and anaerobic filter effluents.

A prior reconnaissance study of anaerobic filter (AF) treatment of pretreated beamhouse wastewaters by Young (2, 3) indicated that up to 50 percent of the biodegradable COD applied to the filter could be removed with a nominal hydraulic detention time of about one day for influent COD values less than 3000 mg/l. Young's work also indicated that inhibition and toxicity problems would occur when AF effluent sulfide concentrations exceeded 200 mg/l. This is important since one of the major components of the keratin hair protein present in tannery wastes is the sulfur containing amino acid cystine. Thus, the anaerobic degradation of hair protein may inadvertently result in the generation of an inhibitory or toxic material that may lead to process failure at high loading conditions. This study, a continuation of Young's work, was designed to systematically evaluate AF performance over a wide range of loading values.

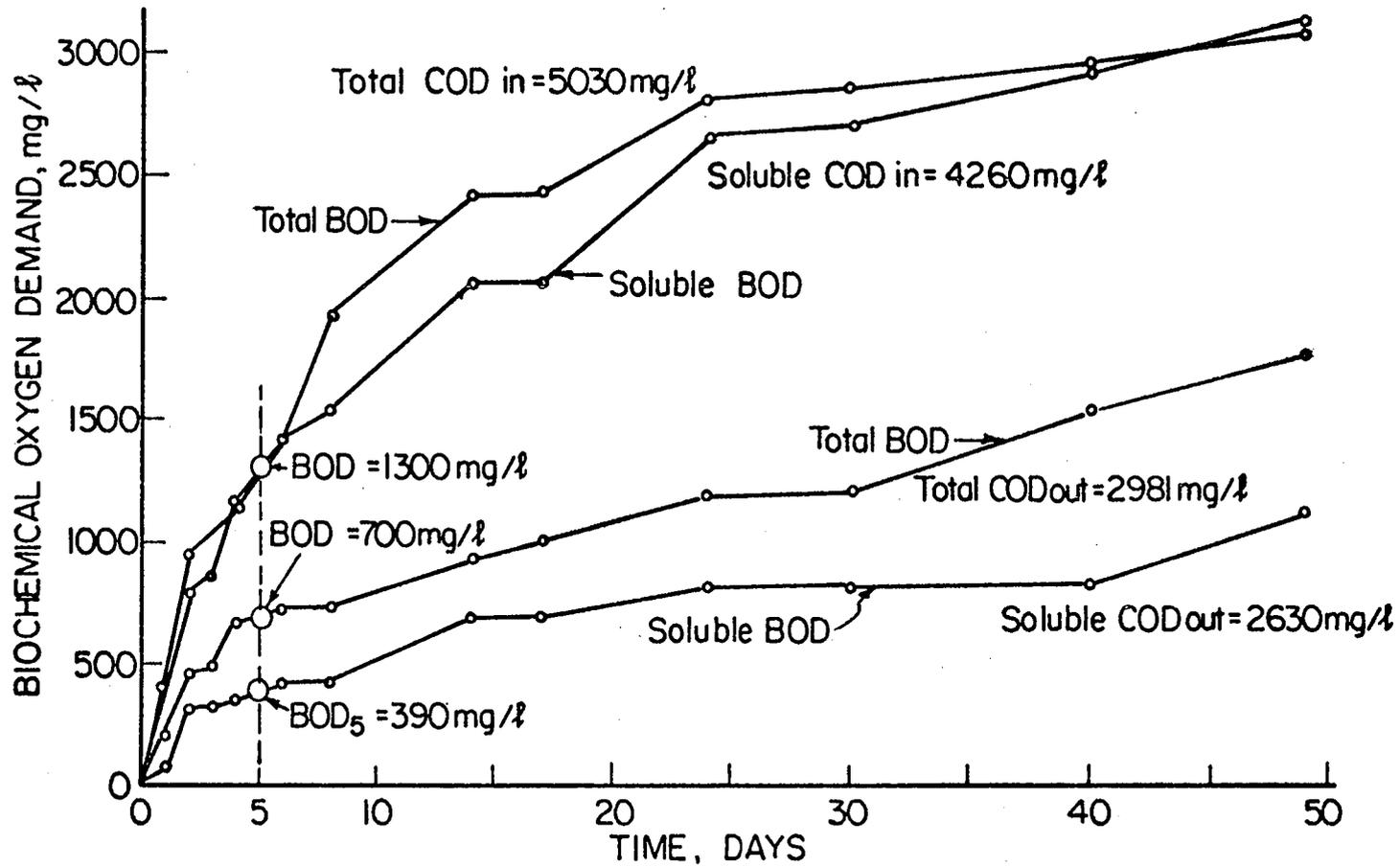


FIG.1 BEAMHOUSE WASTEWATER BIODEGRADABILITY.

EXPERIMENTAL CONDITIONS

Feed

The "standard" beamhouse wastewaters used in this study resulted from the beaming of cowhides and pigskins at the Hides and Leather Laboratory, Eastern Regional Research Center, U.S.D.A. in Philadelphia, PA. Other wastewaters were obtained from the cowhide tanning operations of the Garden State Leather Company in Reading PA. Raw beamhouse wastes were periodically shipped to Syracuse and stored until needed. Previous studies had shown that raw beamhouse wastewaters could be stored in the shipping barrels for extended time periods without degradation due to biological activity. Preparation of a new batch of pretreated beamhouse wastewater consisted of the addition of strong acid (HCl) and mixing until pH 5 was reached and the residual sulfide concentration was less than 300 mg/l. This process both liberated sulfides as H₂S gas and resulted in protein destabilization and flocculation. Following settling for 24 hours, the decant liquor, usually containing less than 150 mg/l sulfides was tested for total and soluble COD and stored at 4°C. Portions of these pretreated beamhouse wastewaters were diluted with tap water to yield the desired COD concentration required for the daily feed preparation.

Reactor System

Each of the two anaerobic filter columns used in these studies consisted of a four foot length of six inch ID clear acrylic tubing divided into seven sections by porous support plates as shown in Figure 2. Each section was filled with Norton Plastic BIO-RING-25 biomass support media* to yield a total clean bed surface area of 3.67 square meters. The clean bed void volume and total volumes were 17.94 and 19.96 liters respectively. Access hatches and sampling ports were located at each lift.

The columns were operated in an upflow mode with feed supplied by a peristaltic pump at a rate of about 18 liters per day, yielding a nominal hydraulic detention time of about one day (Figure 3). In order to prevent plugging, a recycle pump was used to provide a theoretical upflow velocity of about 0.85 feet per minute in the reactor. Several methods were used in attempts to accurately measure gas production.

*Norton Plastics and Synthetics Division, Akron, Ohio

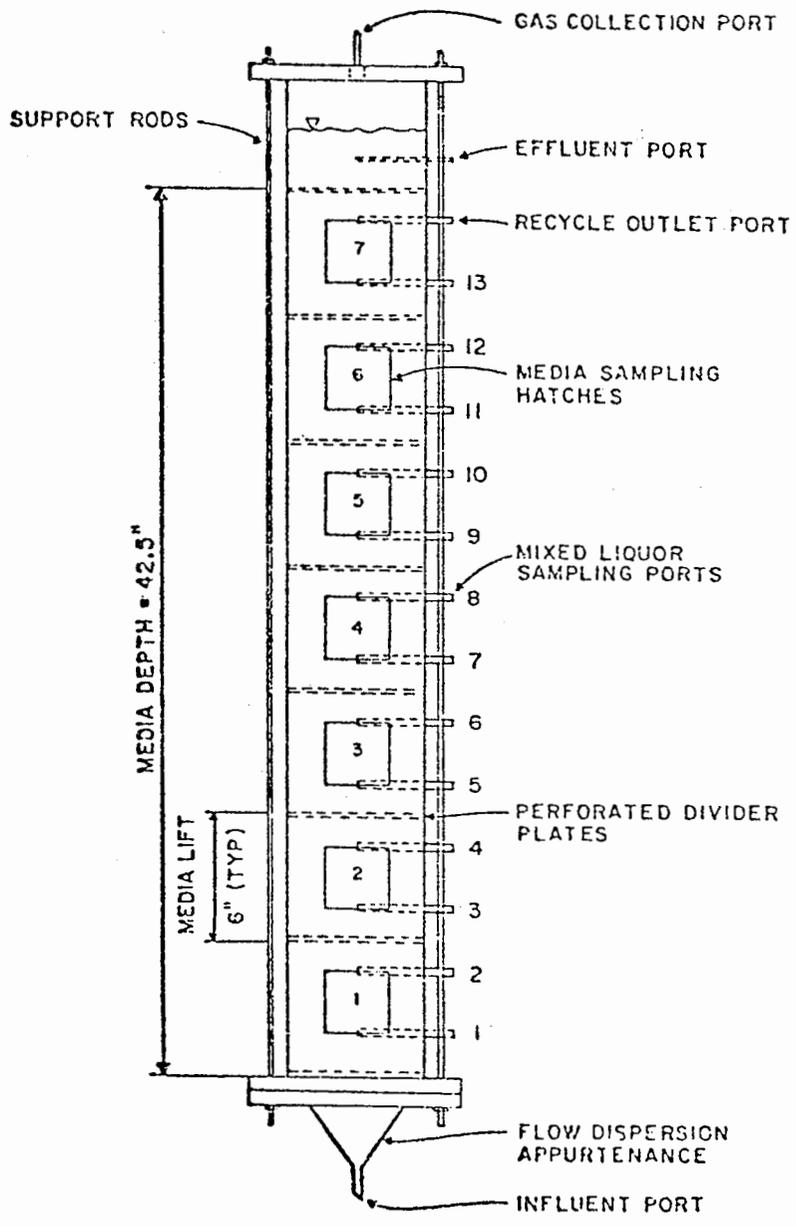


FIG. 2 ANAEROBIC FILTER DETAILS.

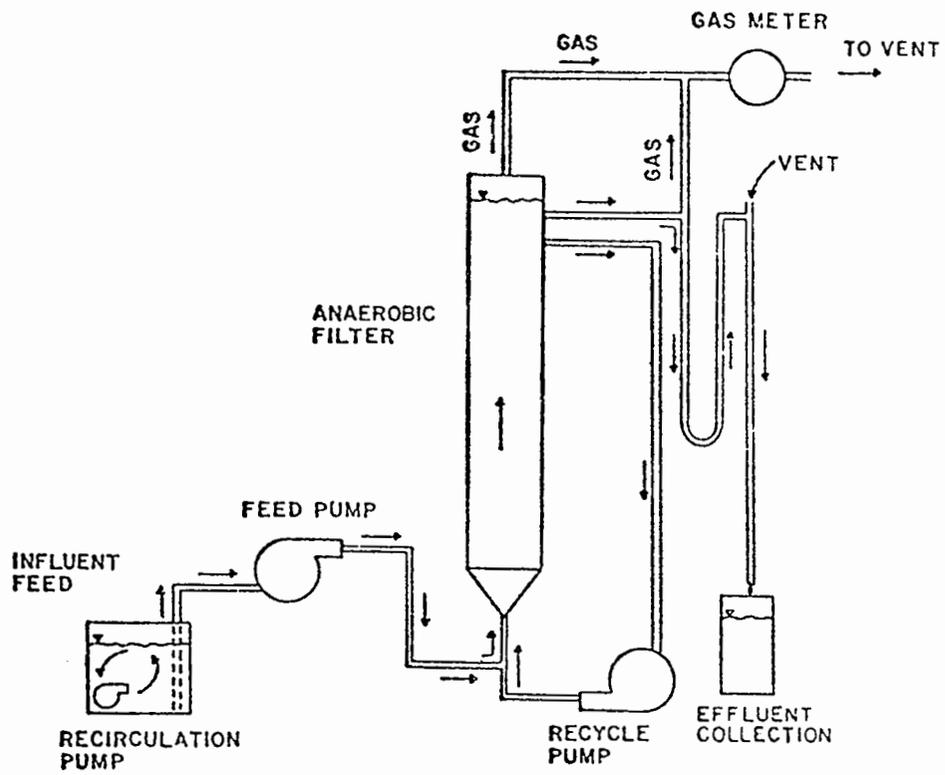


FIG. 3 FEED SYSTEM

Only positive displacement of water proved practical and repeatable.

The two units were housed in a controlled temperature room at $35 \pm 1^\circ\text{C}$ and were originally seeded with anaerobic digester supernatant from a local wastewater treatment plant. Influent and effluent total and soluble COD data, along with feed flow, pH and gas production data were obtained daily. Following establishment of quasi-steady state operating conditions for each loading condition evaluated measurements of alkalinity, acidity, volatile fatty acids, suspended and volatile solids, total solids, gas composition, Folin protein, sulfide and volatile fatty acids were obtained.

RESULTS

Research Difficulties

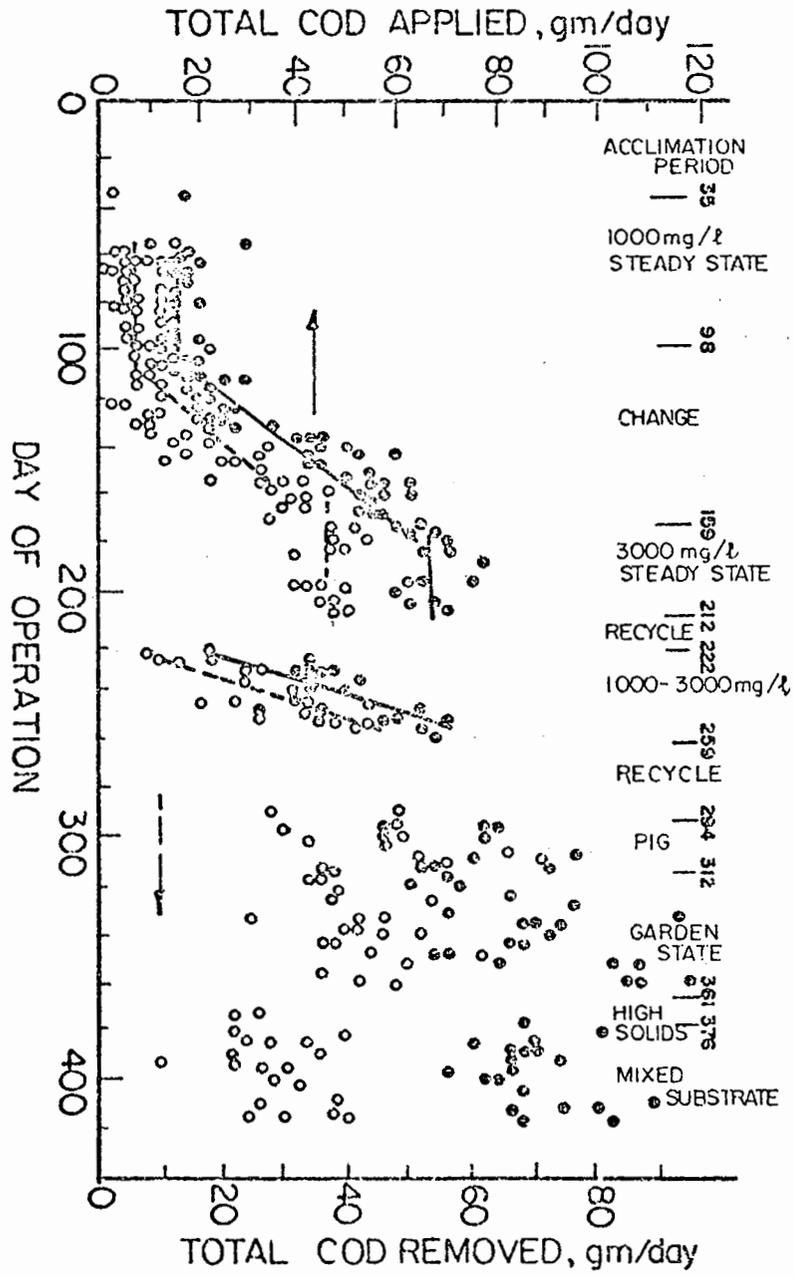
The original research plan was to operate the filters until steady state conditions were observed and then alternately increment organic loadings until process failure occurred in one of the reactors. Unfortunately, a combination of unforeseen mechanical failures and the lack of a consistent supply of "standard" beamhouse wastewater, rather than process failure frustrated these efforts during this 415 day study. Mechanical failures such as power outages and temperature control problems occurred occasionally. Full treatment was generally resumed within a day or two. Experimental delays due to pump problems, parted or clogged feed lines, and reactor leaks were also of relatively short duration.

Of more serious consequence were the problems associated with maintaining an adequate supply of "standard" beamhouse wastewater. During several extended periods it was necessary to reuse previously treated effluent as influent feed. During another period it was necessary to change from cowhide wastewater to pigskin wastewater, resulting in changes of the raw beamhouse wastewater characteristics. Following day 312, raw, "nonstandard" wastewaters obtained from a commercial tannery were used for experimental purposes. During days 361-375 both AF columns were fed a high solids content wastewater that appeared to adversely affect performance. Study loading conditions for AF Column II are indicated in Table I and shown on Figure 4. Similar patterns were observed with Column I. The total COD values shown on Figure 4 have not been corrected for sulfides. Data resulting from mechanical failure or recycle periods have been deleted. Column I failed repeatedly due to biological and physical problems and produced little

TABLE I LOADING CONDITIONS - COLUMN 2

DAYS OF OPERATION	COMMENTS
0 - 35	Startup and acclimation with cowhide wastewaters
35 - 98	Cowhide feed, 1000 mg/l TCOD
99 - 158	Increment feed concentration to 3000 mg/l
159 - 221	Cowhide feed, 3000 mg/l TCOD
212 - 221	Lack of feed, recycled effluent
222 - 258	Cowhide feed, 1000-3000 mg/l TCOD
259 - 294	Lack of feed, recycled effluent
294 - 312	Pigskin wastewater, high grease content
314 - 375	High solids wastewaters
376 - 415	Commercial tannery, mixed feedstock

FIG. 4 COL. 2 - TOTAL COD REMOVAL.



useful data during the latter half of the study period. Although the problems described above prevented the complete attainment of the research goals, they did provide a useful opportunity to study anaerobic filter responses to a number of stress conditions.

Organic Removal Rates

Figure 4 illustrates the rather erratic TCOD loading and removal patterns observed during the latter part of the study. While some of this scatter may have resulted from the use of "nonstandard" beamhouse wastewaters, it is more likely that these data reflect substantially increased areal and volumetric loading rates. At the completion of these studies, the free liquid void volumes were found to have been reduced from a clean bed value of about 17.94 l to 7.83 l and 7.59 l in Columns I and II respectively as a result of solids accumulation. Thus, effective hydraulic detention times were reduced to about 43 percent of the nominal hydraulic detention times and the areal loading rates were substantially increased. While the rate of solids accumulation during different periods of the study is unknown, it is likely that a significant fraction of the accumulation occurred during days 361-375 when high solids feedstock was used. Unfortunately it is not clear whether the scatter in Figure 4 is due to the increased substrate concentrations employed, changes in substrate composition or due to the reduced biomass surface area and exposure time.

When Column II data are arranged as shown in Figure 5, however, a reasonable linear relationship between TCOD application and removal rates appears. A similar relationship was found for soluble COD (Figure 6). Again, the specific cause for the observed scatter could not be ascertained. Similar relationships for Column I at the lower loading levels were observed.

Data from the four "quasi" or "near" steady state loading conditions tested are presented in Tables II and III. Because of pumping and feed problems, the filters did not receive identical loadings from day to day and hence it was impossible to establish true steady state conditions. The terms quasi-steady state or near-steady state as used with the results of this study only imply that a filter was behaving in a consistent manner when subjected to a limited range of loading parameters. Examination of the data in Table II suggest the range of values used for evaluating each "steady state" testing period.

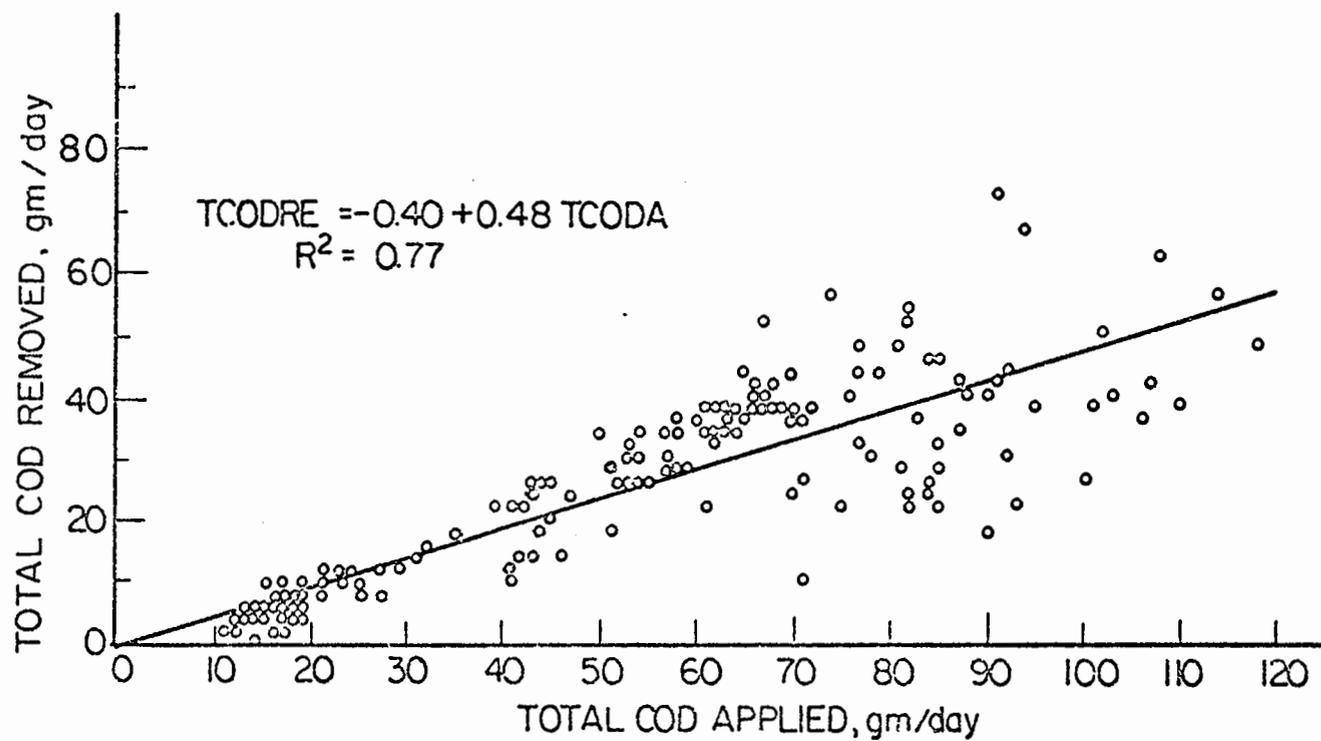


FIG. 5 COL. 2, OVERALL PERFORMANCE-TOTAL COD.

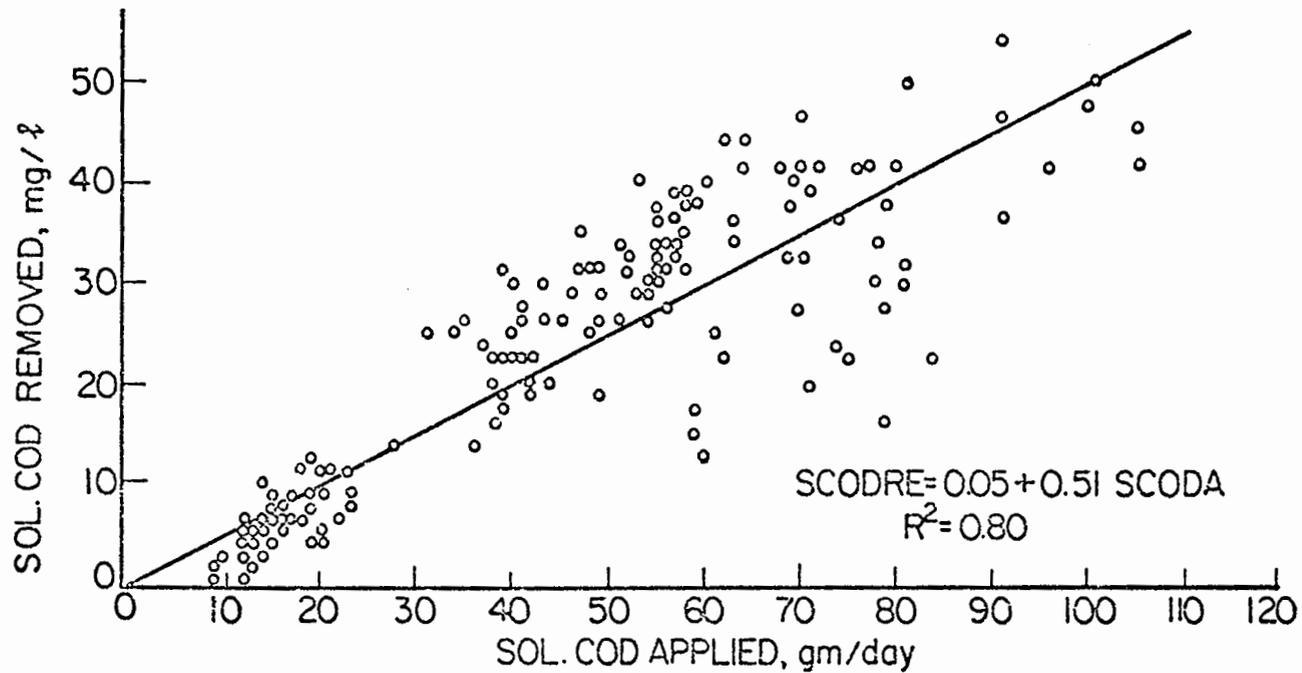


FIG.6 COL. 2 OVERALL PERFORMANCE - SOLUBLE COD.

TABLE II. QUASI-STEADY STATE RESULTS*

TEST PERIOD	I	II	III	IV
Target Influent COD, mg/l	1000	2000	3000	5000
Filter Number	II	I	II	II
Days of Operation	58-92	154-204	294-320	377-415
Influent TCOD, mg/l	886 (173)	2085 (127)	3218 (551)	5088 (558)
Effluent TCOD, mg/l	596 (163)	857 (115)	1156 (246)	3368 (436)
Influent SCOD, mg/l	767 (151)	1661 (137)	2077 (420)	4442 (654)
Effluent SCOD, mg/l	485 (108)	658 (102)	646 (222)	2856 (-)
Flow Rate, l/day	17.6 (1.9)	21.4 (1.5)	22.1 (0.94)	16.4 (1.0)
TCOD Applied, gm/day	15.57 (3.41)	44.69 (4.45)	71.00 (11.51)	83.68 (12.13)
SCOD Applied, gm/day	13.44 (2.59)	35.57 (3.82)	45.97 (9.70)	73.16 (13.48)
TCOD Removed, %	32.5 (12.9)	58.8 (5.6)	63.1 (9.5)	33.6 (6.9)
SCOD Removed, %	35.8 (13.8)	60.3 (5.6)	69.8 (7.1)	35.7 (8.3)
Gas Volume (l/day @ 35°C)	—	—	17.0 (3.9)	10.8 (1.0)

*Numbers in parentheses are the standard deviation

TABLE III TYPICAL DAY

"STEADY" STATE PERIOD	I		II		III		IV	
Day of Operation	91		185		315		413	
Date of Operation	6/05/80		9/27/80		2/04/81		5/13/81	
Feed Flow Rate, l/day	18.6		22.1		21.0		16.8	
Gas Production @ 35°C, l/day	—		—		21.0		16.8	
Column	II		I		II		II	
Parameter	IN	OUT	IN	OUT	IN	OUT	IN	OUT
TCOD, mg/l	804	563	2175	826	2997	1201	6010	3702
SCOD, mg/l	686	420	1636	641	2644	1010	5689	3221
Sol. Polin Proteia, mg/l	385	197	506	65	1093	168	2387	1156
TSS, mg/l	155	83	1352	536	1255	540	537	442
VSS, mg/l	<50	<50	516	183	236	131	325	240
TS, mg/l	3625	3288	7529	5595	17,108	13,853	—	—
VS ₂ , mg/l	623	432	1557	430	2474	1141	—	—
S ₂ , mg/l	15	38	47	73	42	110	68	94 ^e
pH	7.4	7.9	6.4	7.5	6.7	7.5	6.7	7.5
Alkalinity to pH 3.7, mg/l as CaCO ₃	318	554	747	875	435	1135	1013	1745
Acidity to pH 8.3, mg/l as CaCO ₃	32	32	293	115	209	209	367	240
% Methane	—	—	—	—	—	72	—	74

e - estimated

The total chemical oxygen demand (TCOD) mass removal rates were about 0.25, 1.31, 2.24 and 1.41 Kg TCO_D/M³-day on a volumetric basis for testing periods I through IV respectively. Corresponding soluble COD (SCOD) removal rates were 0.24, 1.07, 1.60 and 1.31 Kg SCOD/m³-day. The relatively low COD removal rates observed for the first testing period were probably due to the lack of sufficient time to develop a mature biomass population in the column. The COD removal rates observed during the fourth testing period may reflect the effects of either reduced detention times in the column or adverse concentration effects.

The data in Table III, provide interesting insight to the dynamic conditions occurring within the column. As keratin protein is degraded, deamination reactions result in the generation of ammonium ions and concurrent increases in pH and alkalinity. Similarly, as the disulfide bond in the keratin is destroyed, sulfides are produced. While both total and volatile solids were being removed within the columns, it is not clear that they were being entirely destroyed by biological activity. By the end of the study about 10 liters of solids had accumulated in each column despite a massive recycle ratio and internal mixing due to gas production. Samples extracted from the bottom, midpoint and top of the filters on each of the four days shown in Table III, yielded significantly different values for each of the parameters. It is clear that these columns could not be modeled as simple complete mix reactors.

Figure 7 illustrates TCOD removal performance during the first 313 days of the study with Column 2 and USDA "standard" beamhouse wastewaters. These data reflect widely varying loading conditions with influent concentrations ranging from about 600 mg/l to 3500 mg/l TCOD and flow rates ranging from 13 to 26 liters per day. Data obtained during recycle periods and days of feed pumping problems have been deleted. Both pretreated cowhide and pigskin beamhouse wastewater treatment data are intermixed and are indistinguishable from each other. These data also indicate that prolonged feed stoppages of up to forty days had no effect on subsequent performance. Based on these observations, it appears that a mature anaerobic filter system operating with pretreated beamhouse wastewaters is relatively insensitive to fluctuations in daily loading conditions up to 2.25 Kg TCO_D/M³-day and is able to recover quickly from a wide variety of environmental stresses.

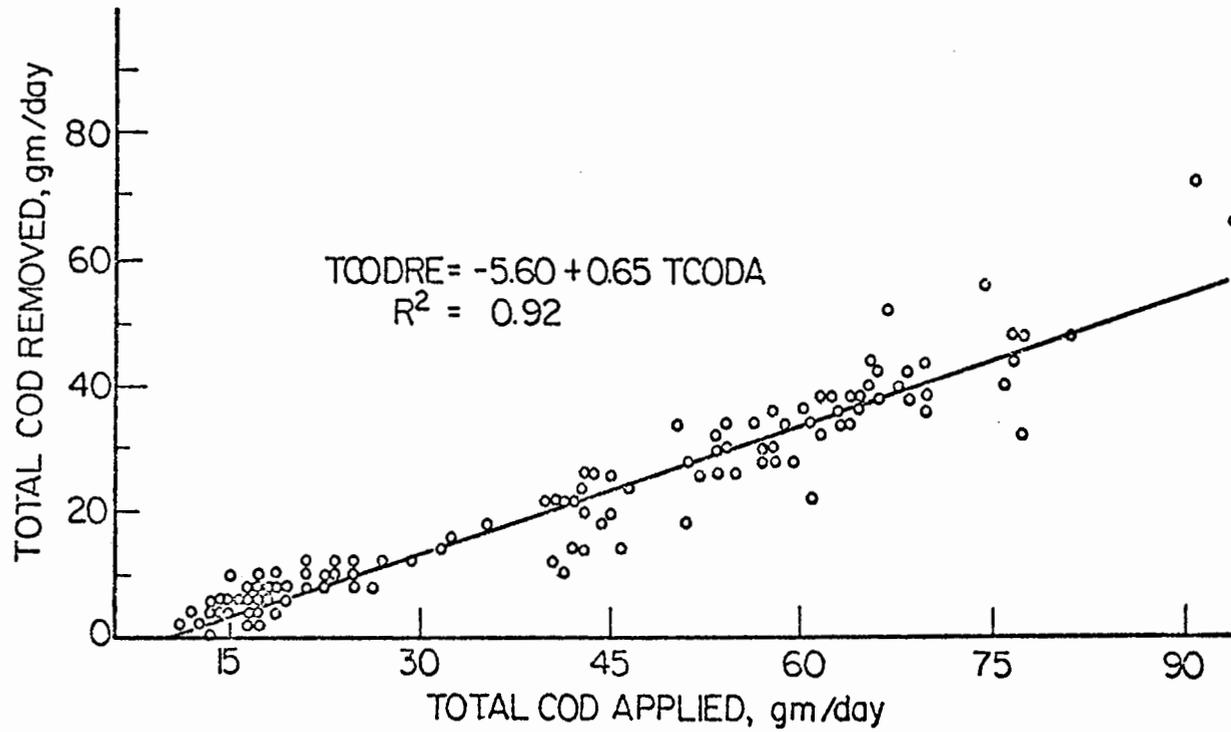


FIG. 7 USDA BEAMHOUSE WASTEWATERS.

Gas Yield

Methane production is a desirable benefit of anaerobic systems. Figure 8 illustrates that the average gas yield for this system remained reasonably constant despite the variety of feed sources and loading rates employed from day 222 through day 415. The average gas yield value was found to be about 0.48 liters of total gas at 35°C per gram of TCOD removed. Since methane constituted about 74 percent of the gas collected, the methane yield averaged 0.31 l/gram TCOD removed at STP. It was evident that objectionable quantities of H₂S were generated in the gas stream. Gas scrubbing systems will have to be an integral part of any anaerobic beamhouse wastewater treatment system.

Solids Accumulation

As previously mentioned, upon completion of the study the filters were drained and the void volumes were found to be about 43 percent of the original void volume. When the access hatches to each lift were opened, it was apparent that all surfaces were coated with dark gelatinous layers up to one centimeter thick. All of the BIO-RINGS examined were at least partially plugged. The dark materials contained gritty sand-like particles that would neither dissolve in strong acid or base. The volatile fraction of these solids increased from 19.8 percent at the bottom of Filter II to 25.2 percent at the top. About a third of the nonvolatile matter was calcium, which was not surprising considering that the raw beamhouse wastewaters were saturated with lime. This massive deposition of solids suggests that anaerobic filter reactors treating wastes that are likely to form precipitates should be operated in a downflow mode to aid in washing solids from the system. Anaerobic filter system designs should include provisions for periodically backwashing or surging the media to help loosen excess solids.

Aerobic Treatment

Effluent from the anaerobic filters was subjected to further aerobic biological treatment with bench scale rotating biological contactor (RBC) units. The RBC study results will be reported shortly (4) and will provide data suggesting that the anaerobic roughing treatment of beamhouse wastes removes protein constituents that would otherwise have escaped treatment in the aerobic system.

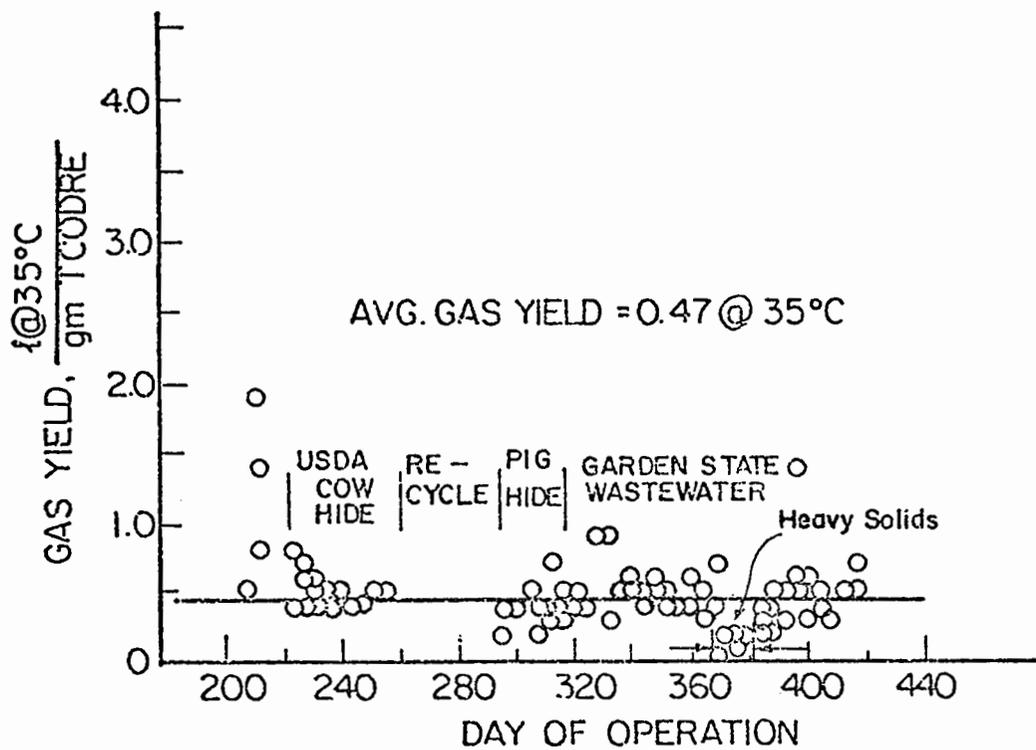


FIG. 8 GAS YIELD.

CONCLUSIONS

These studies have shown that pretreated tannery beamhouse wastewaters can be treated with an anaerobic process to routinely remove more than 60 percent of the applied TCOD with hydraulic detention times less than one day and loadings on the order of 2.25 Kg TCOD/m³-day. The lack of noticeable production of volatile fatty acids and pH depression across the columns for these loading conditions suggests that the rate limiting step for this anaerobic process is the degradation of proteins into peptides and amino acids. The anaerobic system proved to be remarkably insensitive to variations in feed type, loading conditions, shut down periods and the introduction of air into the system.

The massive accumulation of solids within the reactors suggests the employment of larger support media than that used in this study and that a down flow mode of operation be used for systems where reaction conditions may result in precipitate formation in order to minimize the potential for plugging and solids accumulation. Gas collection systems for methane recovery will have to include provision for sulfide scrubbing.

ACKNOWLEDGEMENTS

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ANAEROBIC FLUIDIZED BED TREATMENT OF WHEY:
EFFECT OF ORGANIC LOADING RATE, TEMPERATURE AND
SUBSTRATE CONCENTRATION

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ABSTRACT

The anaerobic biological fluidized bed process has been shown to be an extremely efficient system for the treatment of moderate and high strength industrial wastes. This is due to the extremely high specific surface and mass transfer properties of the fluidized bed system. Reported in this article are the results of over three years of pilot work on generating methane gas from whey using a Hy-FloTM anaerobic fluidized bed. The effect of organic loading rate, influent substrate concentration and temperature has been examined. Energy balances show the process to be a net energy producer, generating between 30.5 and 78.7 times the amount of energy required to operate the system.

INTRODUCTION

Whey is a slightly viscous, greenish-yellow liquid by-product of cheese manufacture. It is extremely rich in

lactose (5.0 percent) and protein (0.9 percent). Reported BOD₅ values range from 30,000 to 40,000 mg/l (1-4). Total solids and COD values normally vary between 6.0 and 6.5 percent. The economical disposal of this high strength organic waste constitutes a major problem for the cheese industry. Condensing and drying whey, once thought to be the optimal method of disposal, is rapidly losing popularity due to the energy intensive nature of the condensing and drying operations and lack of a stable market for the lactose. Due to its high strength, whey is an excellent candidate for energy recovery. The two options currently under consideration for this purpose are generation of methane gas and alcohol production.

A recent survey of New York State cheese manufacturers reported that the energy needs of a cheese plant could be reduced by an average of 35 percent through anaerobic methane fermentation of whey and the subsequent use of the methane on site (5). Energy balances show the process to be a large net energy producer (6). Alcohol production by contrast, currently has a net negative energy balance due to the dilute nature of the distillation feedstock (7). A 40 cent per gallon federal subsidy is the main factor that allows alcohol production to remain competitive with methane generation. The generation of methane gas offers several other significant advantages. The process is relatively simple and less capital intensive than alcohol production and the technology can be used for any cheese plant regardless of size. Alcohol production becomes attractive only for extremely large volumes of whey. Normally this would require the central processing of whey from locally competitive dairies (7).

METHANE GENERATION FROM WHEY

Fermenting whey to methane gas was initially investigated by Buswell (1). In a conventional CSTR laboratory digester 93 percent BOD₅ reduction was obtained at a loading rate of 2.2 gr of volatile solids per liter of digester per day. More recently, Parker (4) attained 99 percent BOD reduction in conventional 500 gallon pilot digester system with a 6-7 day retention time. The maximum organic loading

rate was reported to be 4.3 kg BOD/m³/day. Danskin (8) using diluted whey (1 percent solids) was able to achieve 93.1, 92.8 and 87.7 percent COD reduction at organic loading rates of 8.9, 13.3 and 20.0 kg COD/m³/day respectively for a bench scale fluidized bed system. The hydraulic retention time ranged from 12 to 27 hours. The study detailed herein reports the results obtained using a pilot scale anaerobic fluidized bed to treat acid whey. Loading curves for whole whey (6 percent solids) at 35°C and 24°C and diluted whey (1 percent solids) were generated. Results indicate that whole and dilute whey can easily be treated at a high rate in an anaerobic fluidized bed and that this system offers greatly reduced reactor sizes compared to alternate anaerobic treatment systems.

MATERIALS AND METHODS

Testing was conducted with standard skid mounted Hy-FloTM fluidized bed pilot (Figure 1). The pilots consist of a nominal 6-inch diameter by 10-foot high clear PVC bioreactor equipped with gas separation and measurement chambers, temperature controller and the necessary feed, recycle and chemical addition pumps. Temperature was normally controlled at 35°C and pH maintained in the 6.8-7.4 range by adding sodium bicarbonate directly to the influent wastewater.

To avoid storage problems, whey powder was used as the feedstock. Fresh influent was prepared on a daily basis. Daily analysis for pH, alkalinity, volatile acids, total and soluble COD were performed on influent and effluent samples. Information on feed rate, temperature and fluidized bed volume was recorded daily. Routine monitoring of total and soluble BOD₅, suspended solids and gas production and composition were also performed. All tests with the exception of volatile acids were performed in accordance with Standard Methods (9). Volatile acids were analyzed by the method proposed by O'Brien and Donlan (10).

In order to reduce the time required for start-up, the reactor was initially charged with sand media used in previous studies instead of virgin sand. Initially, the reactor was loaded at roughly 1.0 kg COD/m³/day. This was increased as rapidly as possible to the first loading rate of interest. The organic loading rate was then maintained

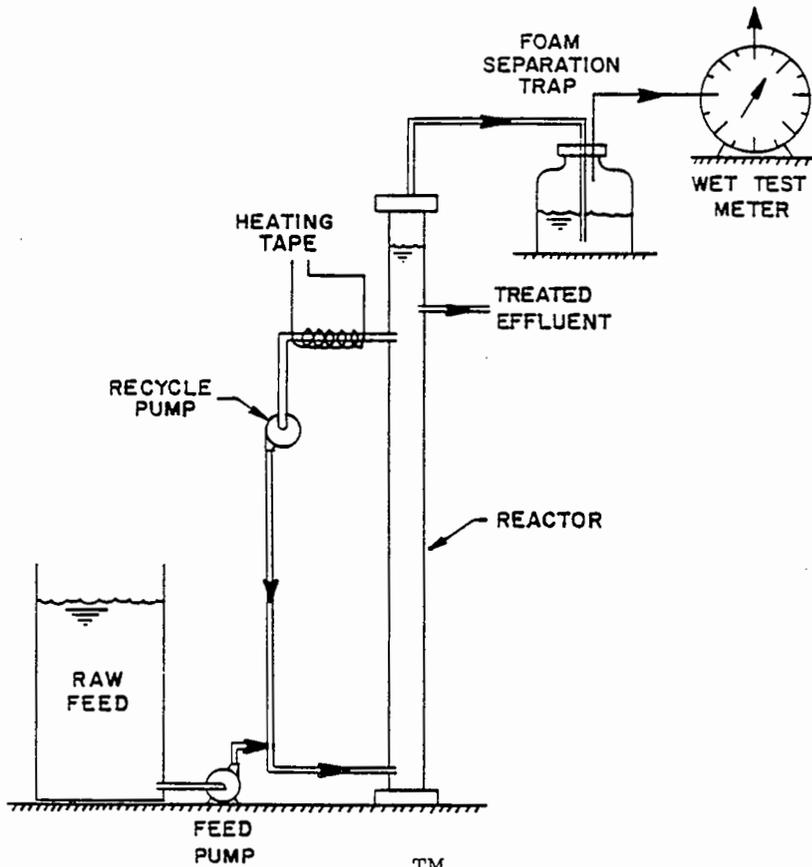


Figure 1. Schematic of Hy-FloTM Fluidized Bed Pilot.

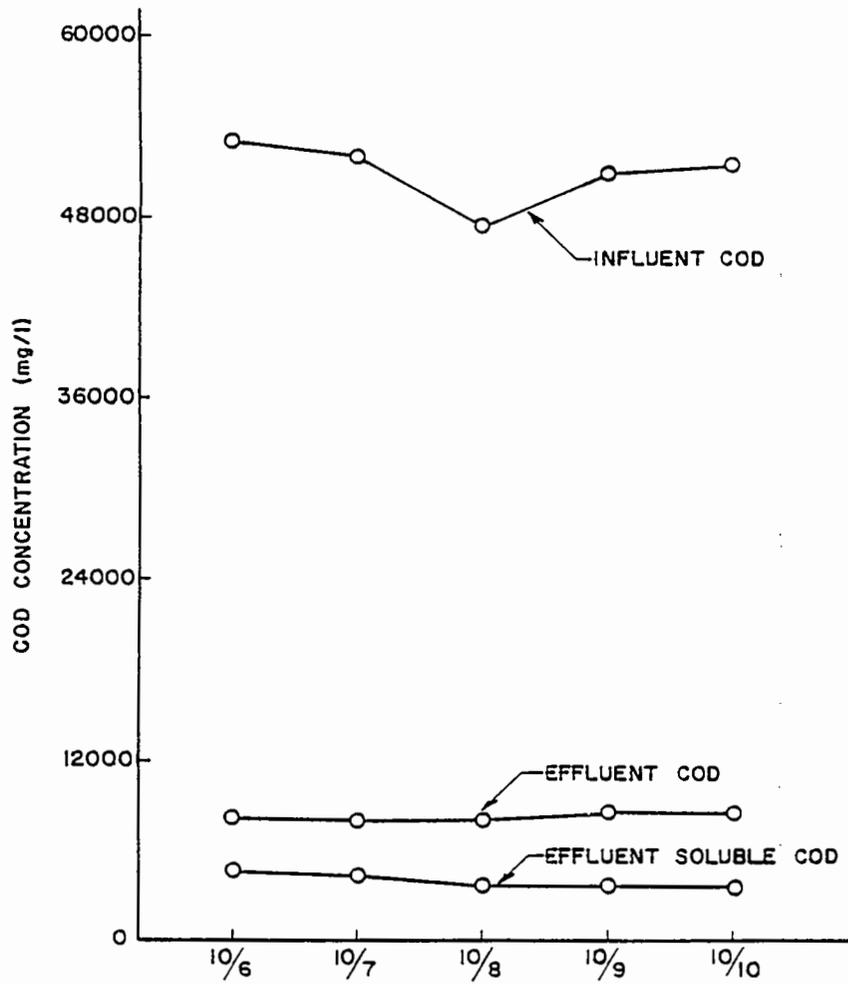


Figure 2. Typical Data for Reported Steady State Data Points used in Loading Curves.

relatively constant until reactor performance stabilized. Normally this required a minimum of 3-5 hydraulic residence times. Collection of one or two weeks' steady state data was considered sufficient to assess system capabilities. The loading rate would then be increased or decreased to the next level of interest and the above procedure repeated. Typical steady state operating data is shown in Figure 2.

RESULTS

The study was divided into three main phases. Initially, diluted whey powder (1 percent solids) was utilized as the feedstock. This simulates the combined treatment of whey and cheese plant process wastewater. Following collection of sufficient steady-state information to generate a loading curve (percent COD reduction versus applied organic loading rate) the influent substrate concentration was increased to that of "whole whey" (6 percent solids). A loading curve for whole whey was then generated. The reactor temperature was held constant at approximately 35°C throughout both the initial two phases of the study. While continuing to feed the whole whey to the reactor, the temperature was dropped to 24°C for phase three. Sufficient time was allowed for the reactor to adjust to the new operating conditions (between 8 and 9 hydraulic retention times) and a three week run was made. The loading rate was readjusted and a two week run was made following a period of approximately 10 retention times. Since whey normally leaves the cheese plant at elevated temperatures (greater than 30°C), the 24°C operating temperature could be considered a "worst case" where no supplemental heating is provided to maintain the reactor temperature.

Effect of Organic Loading Rate on Removal Efficiency - 1 Percent Solids

Eight runs of a one two week average duration were made during the initial phase of operation. Organic loading rates of between 4.0 and 32.4 kg COD/m³/day were examined and are summarized in Table 1. The hydraulic retention time (based

TABLE I. Summary of Results Using 1 Percent Solids

Run No.	COD (mg/l)		Percent Removal	pH	Volatile Acids (mg/l as HAc)	Organic Load kg COD/m ³ /day	GAS PRODUCTION	
	Influent	Effluent					$\frac{1 \text{ CH}_4}{\text{gr COD}_r}$	HRT* (hrs)
1	10,210	2,870	71.9	6.9-7.3	970	8.2	0.35	22.9
2	9,750	4,190	57.0	6.9-7.2	1,530	16.4	0.30	14.6
3	9,730	4,540	53.3	7.0	1,950	19.6	0.36	12.0
4	10,070	3,690	63.4	7.0-7.3	1,575	18.1	0.36	13.2
5	10,760	4,890	54.6	6.8-7.2	2,430	32.4	0.40	8.0
6	11,060	1,550	86.0	7.0-7.2	250	6.1	0.41	44.5
7	9,170	1,020	88.9	7.0-7.1	160	4.0	--	55.6
8	9,025	950	89.5	7.0-7.2	180	4.5	--	47.6

*Empty Bed Retention Time

on empty bed volume) ranged from 8.0 to 55.0 hours. COD removals were found to vary from a high of 89.5 to a low of 53.3 percent. BOD removals were slightly higher ranging from 96 to 53.8 percent.

Results for COD removal versus organic loading rate is presented in Figure 3. Despite the fact that the fluidized bed showed excellent removal capabilities, it is quite possible that the unit had not totally acclimated to the whey. This observation is based on the fact that runs made at the end of this segment of the study showed much higher removals than would have been predicted by an extension of the partial loading curve obtained from the initial information (Fig. 3).

Effect of Organic Loading Rate on Removal Efficiency - 6 Percent Solids

Nine runs were completed using "whole whey" as the feedstock for the fluidized bed pilot. The initial seven runs were made at 35°C and the subsequent two runs at 24°C. Results are summarized in Table II and Figure 4. Applied organic loading rates of between 13.4 and 37.6 kg COD/m³/day were examined at 35°C with resulting COD removals of 83.6 to 72.0 percent. This represents hydraulic retention times of between 1.4 and 4.9 days. Removals were quite stable over the entire range of loading rates and appeared to drop almost linearly with increasing loading rates.

The results at 24°C were also impressive. At organic loading rates of 15.0 and 36.8 kg COD/m³/day, COD removals averaged 71.0 and 65.2 percent respectively.

Series Operation

Near the end of phase II (35°C and whole whey) a week of steady state operation with reactors in series was obtained, by using another pilot to treat the effluent from reactor #1. Results are summarized in Table III and Figure 5. The first reactor in the treatment train was operated as a roughing unit (loaded at 37.6 kg COD/m³/day). Reactor #1 removed 73 percent of the BOD₅ (72 percent of the COD) while reactor #2 removed 87 percent of the residual BOD₅ (78

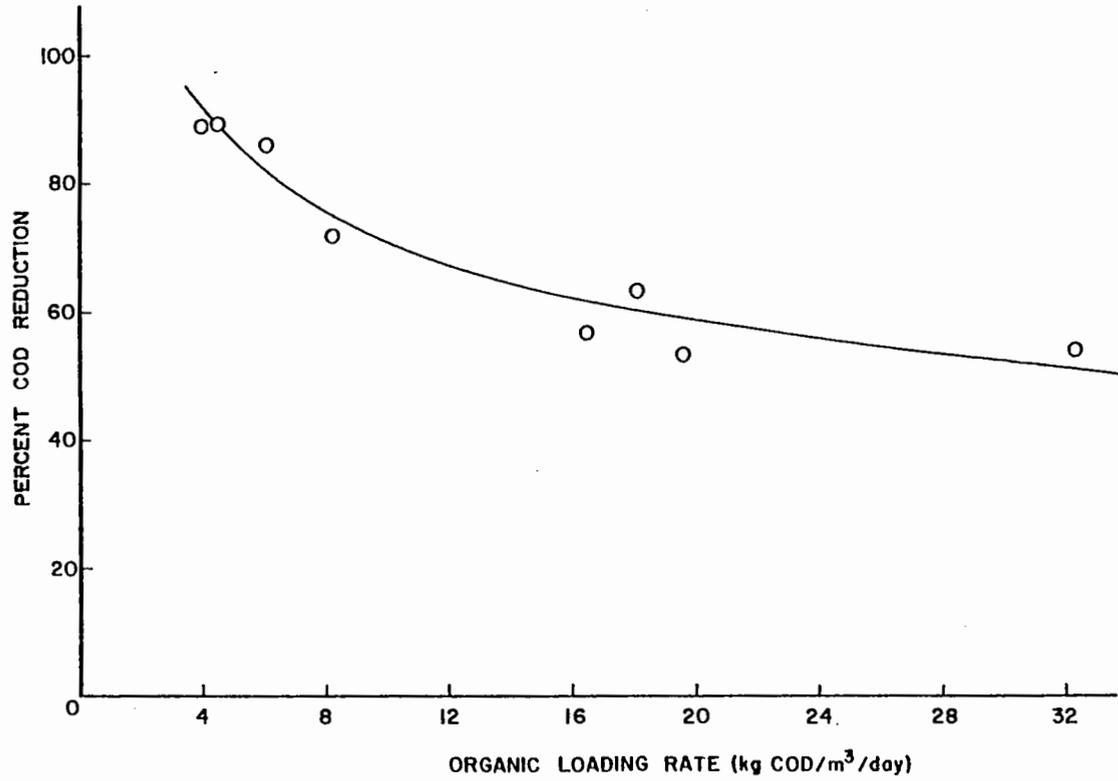


Figure 3. Effect of Organic Loading on COD Removal for 1 Percent Whey.

TABLE II. Summary of Treatment of Whole Whey

Run No.	Temperature °C	COD's			pH	Volatile Acid (mg/l) as HAc	HRT* (days)	Organic Loading Rate (kg COD/m ³ /day)
		Influent (mg/l)	Effluent (mg/l)	Removal %				
1	35	55510	14500	73.9	6.5-7.1	3510	1.5	37.1
2	35	51540	13325	74.1	6.9-7.1	2160	1.75	29.6
3	35	56090	12500	77.7	6.9-7.3	1800	2.0	28.3
4	35	53490	12820	76.9	7.0	2000	2.15	25.8
5	35	52260	14590	72.0	6.8-7.0	3630	1.4	37.6
6	35	50320	9640	80.8	7.0-7.2	2150	2.9	17.3
7	35	50890	8350	83.6	7.1-7.3	1360	4.9	13.4
8	24	52240	15170	71.0	6.9-7.1	1270	3.5	15.0
9	24	55410	19275	65.2	6.8-7.0	3000	1.5	36.8

*Empty Bed detention Time.

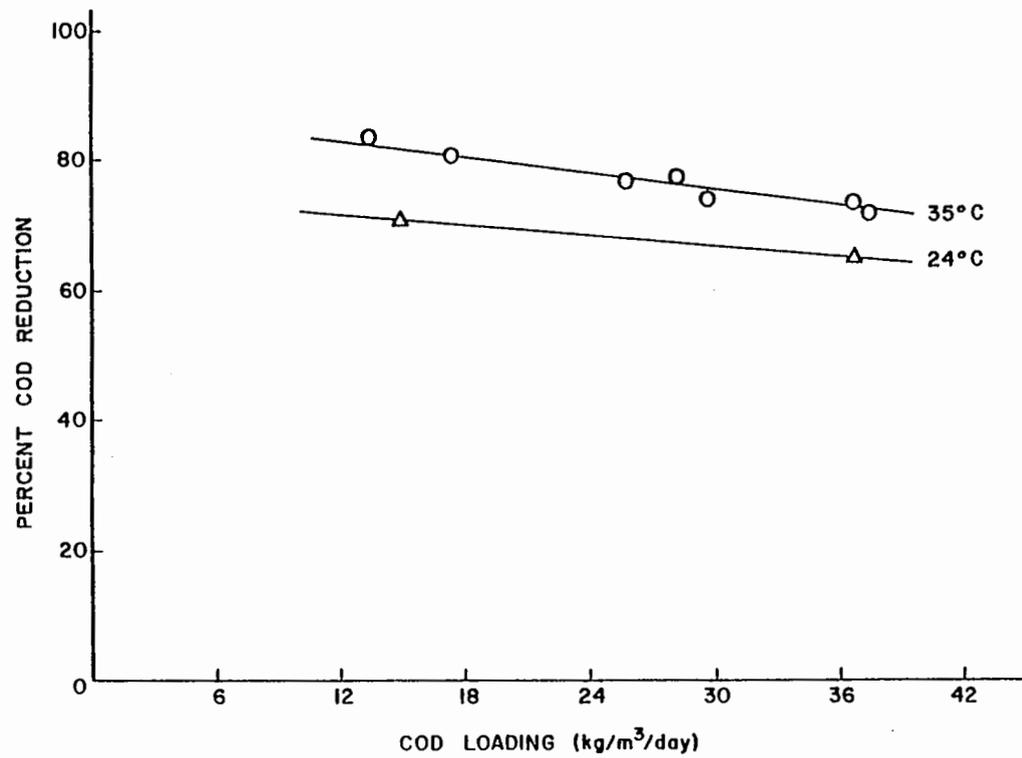


Figure 4. Effect of Organic Loading Rate on COD Removal Rate for Acid Whey.

TABLE III. Series Operation - Organic Loading Rates

	Organic Loading ₃ (kg COD/m ³ /day)	Percent COD Removal	HRT* (Days)	% BOD ₅ Removal**
Reactor 1	37.6	72	1.4	73
Reactor 2	2.7	78	3.6	87
Reactor 3	10.5	94	5.0	96.5
		(95.5)***		(97)***

*Based on Empty Bed Retention Time

**Only one observation

***Total Influent to Soluble Effluent

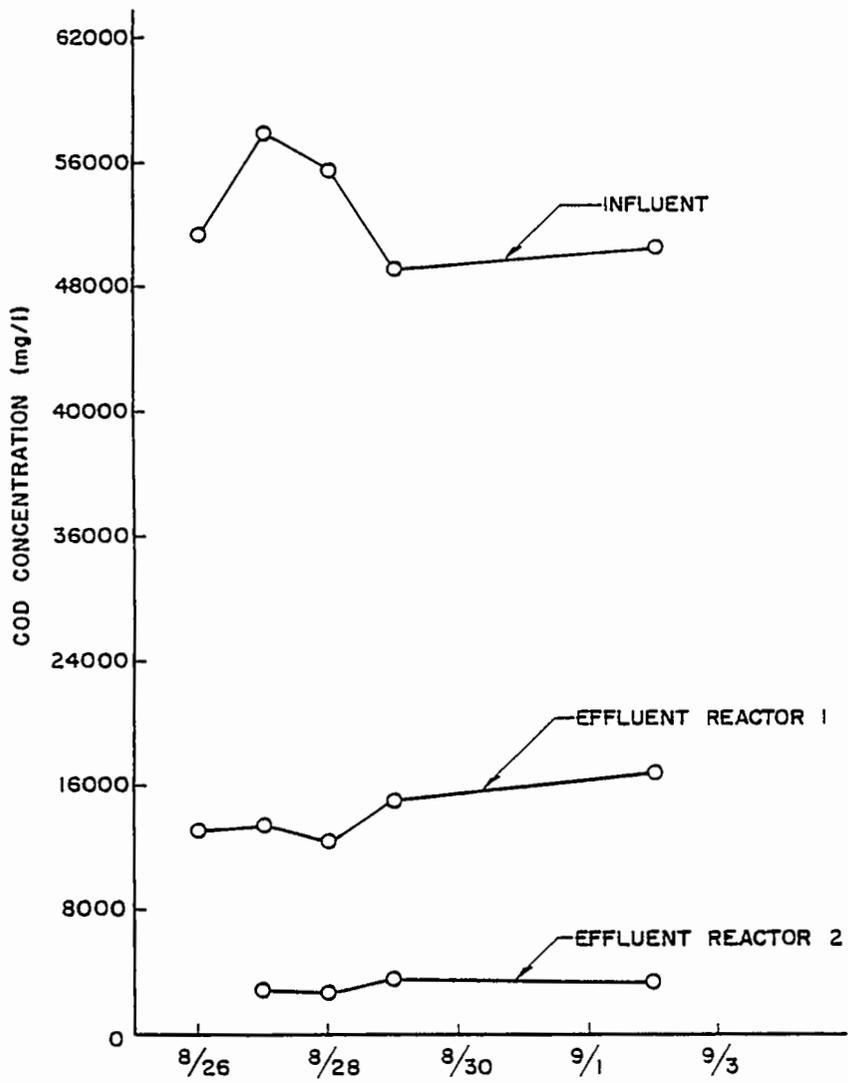


Figure 5. Daily COD Results for Reactors in Series Treating Acid Whey.

percent COD), for an overall BOD₅ removal efficiency of 96.5 percent (94 percent COD). The combined hydraulic retention time was 5.0 days (1.4 days in reactor #1 and 3.6 days in reactor #2) which converts to an overall organic loading rate of 10.5 kg COD/m³/day. Approximately 85 percent COD removal would be predicted with only one reactor at this loading. Although the series operation appears to be quite effective, there are other considerations which must be taken into account. The extremely high organic loading imposed on reactor #1 resulted in high volatile acids (3950 mg/l as HAC) which requires higher alkalinity additions to maintain the reactor pH in a favorable range. For optimal design, a tradeoff of capital versus operating cost must be made.

Effect of Organic Loading Rate on Volatile Acids Production

The fluidized bed reactor volatile acids concentration was found to be a strong function of the organic loading rate (kg COD/m³/day). These results are depicted graphically in Figure 6. Both 1 percent solids and 6 percent solids are displayed together. It can be seen that volatile acids concentration increases sharply with increasing loading rates. Despite the high volatile acids concentrations encountered at the higher end of the loading curves (volatile acids up to 3630 mg/l as HAC), no inhibition in reactor performance was apparent.

Gas Production

Methane content of the gas was assumed to equal total gas production minus the fraction attributed to CO₂. The percentage of CO₂ was monitored using a Bachrach² Apparatus, which is capable of measuring CO₂ only to within several percent.

The off gas average approximately 60 percent methane and 40 percent CO₂. The methane production averaged 0.363 liters of CH₄ per gram of COD removed. This is 92 percent of the theoretical value of 0.395 liter of CH₄/gr COD predicted by stoichiometry.

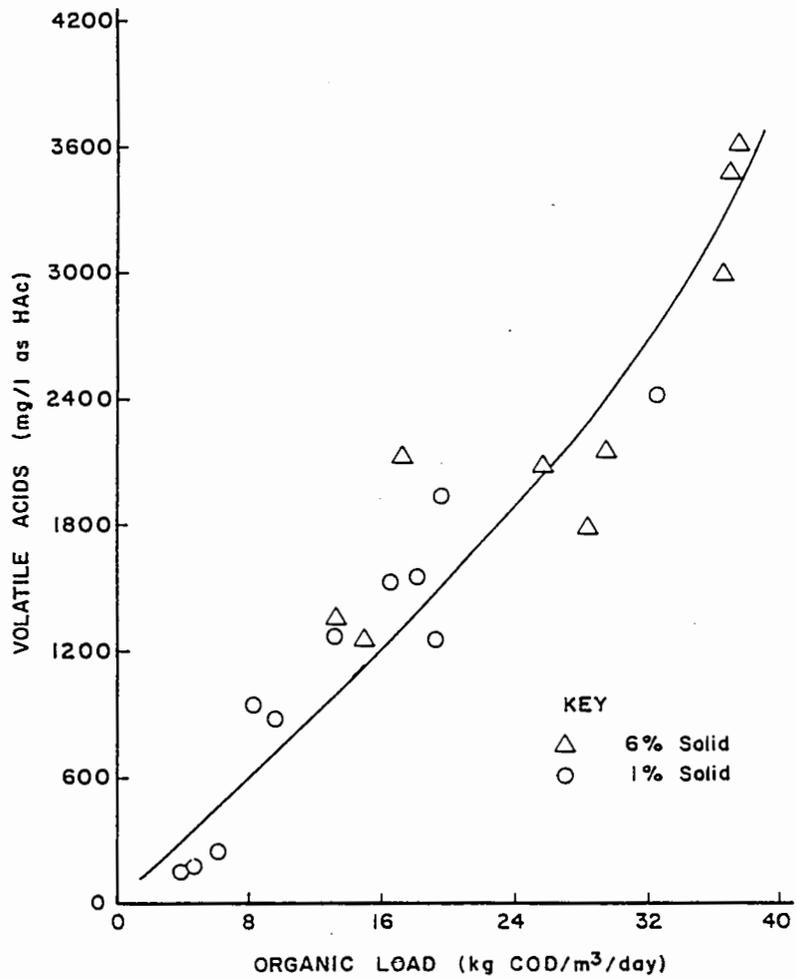


Figure 6. Effect of Organic Loading Rate on Volatile Acids Accumulation for Whey Treatment.

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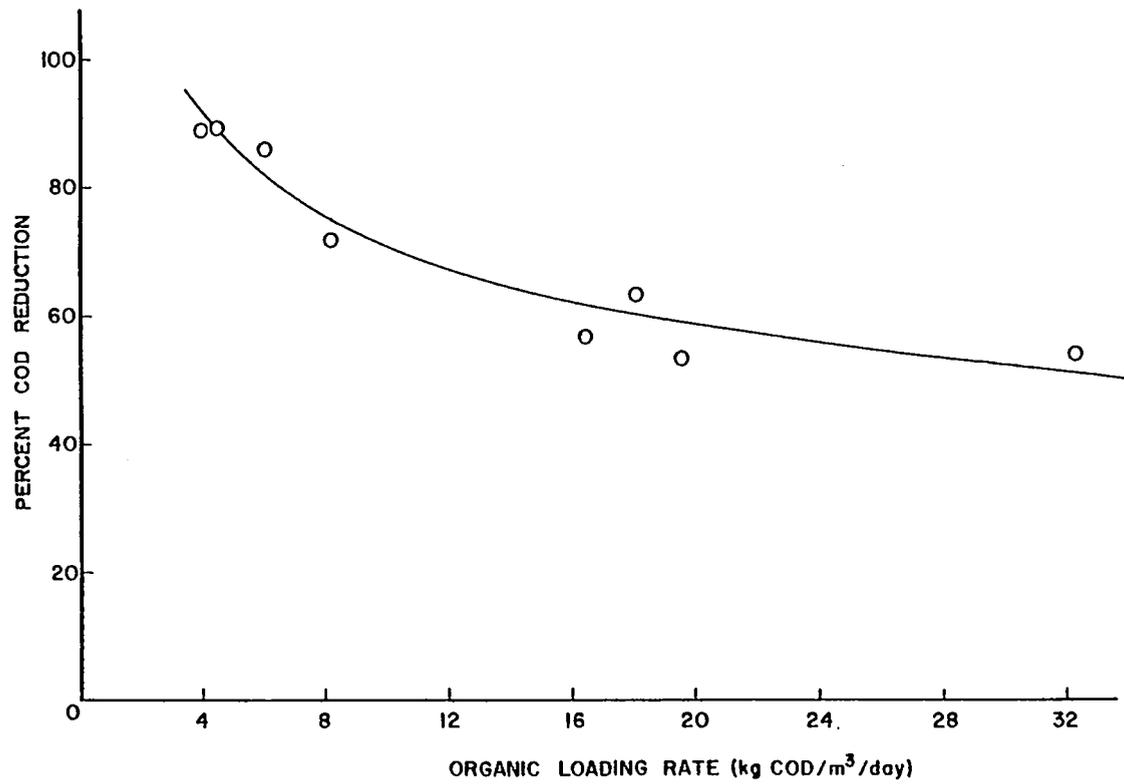


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5	35	52260	14590	72.0	6.8-7.0	3630	1.4	37.6
6	35	50320	9640	80.8	7.0-7.2	2150	2.9	17.3
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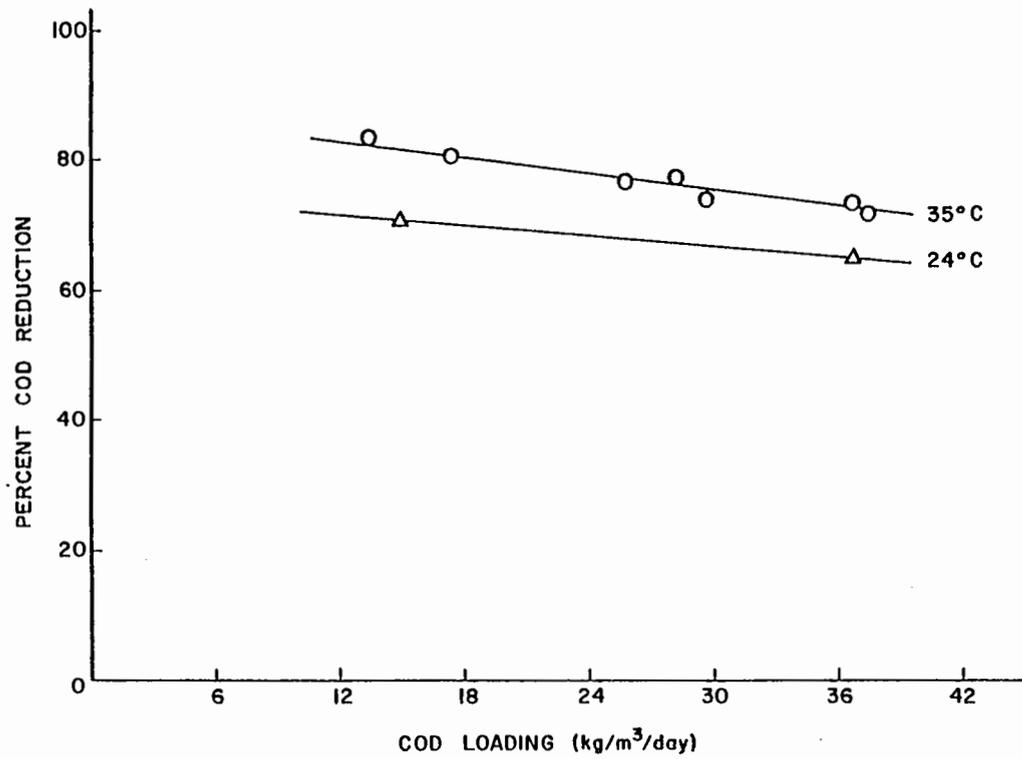


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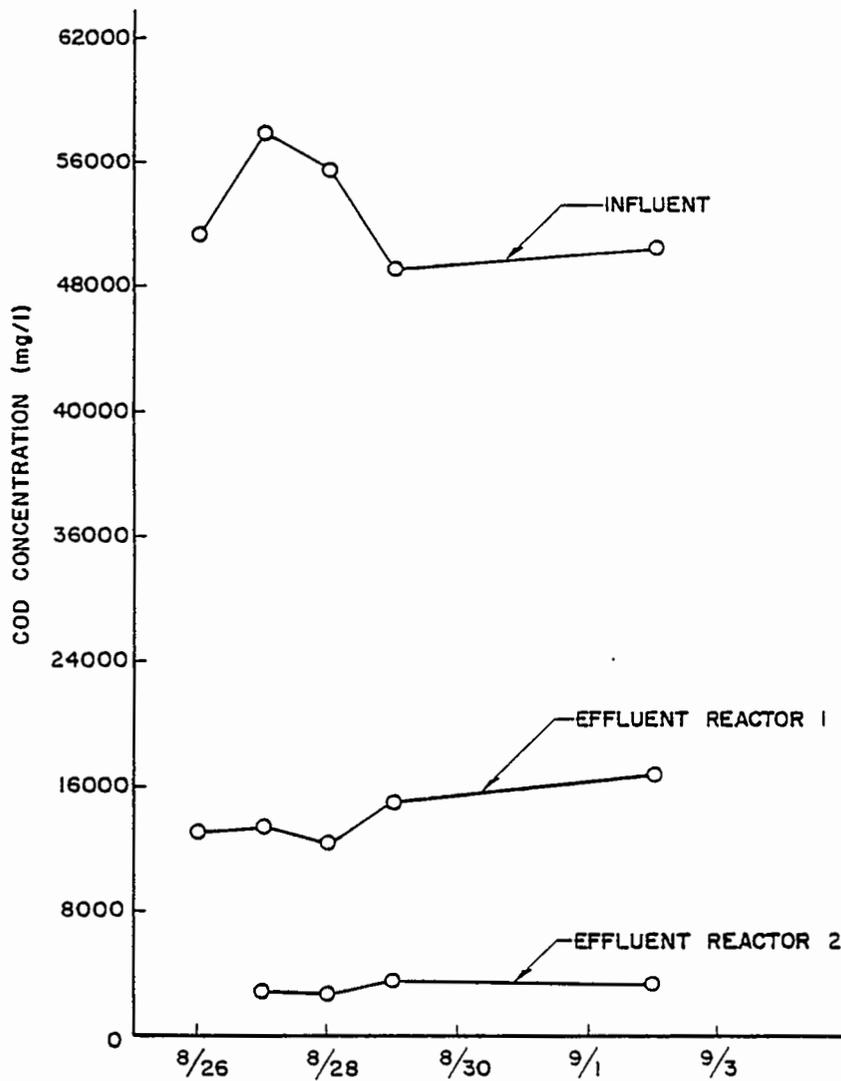


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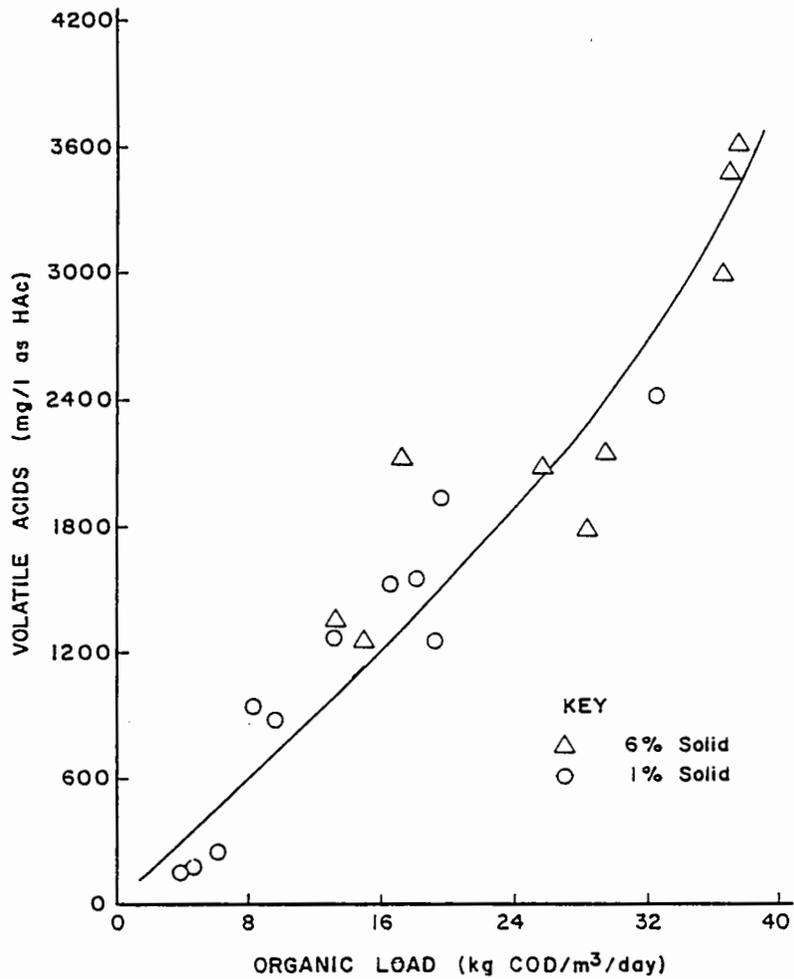


Figure 6. Effect of Organic Loading Rate on Volatile Acids Accumulation for Whey Treatment.

ENERGY BALANCE

The energy balance of treating the whey from a cheese plant producing 170,000 pounds of fluid whey per day was calculated for five loading rates. Methane production was estimated assuming a yield equal to that observed during the pilot testing, 92 percent of the theoretical, and further assuming that 95 percent of that gas could be recovered and utilized. The methane gas energy value was set at 1000 BTU's per standard cubic foot. Energy usage was calculated based on the estimated running horsepower of the influent, chemical and recycle pumps and chemical mixing equipment.

As is evident from Table IV, a large amount of energy is generated. Net production is a maximum at the lowest applied organic loading rate (15.07×10^6 kwh/yr) and lowest at the highest loading rate (13.21×10^6 kwh/yr). The energy yield ratio, that is the ratio of energy produced to energy consumed, shows the opposite trend running from 30.5 to 78.7 for the lowest to the highest loading rates. In all cases, the anaerobic fluidized bed system shows remarkable energy production. If the loading curve developed in the pilot study were to be extended to cover lower applied organic loading rates, the maximum net energy production would occur close to $12 \text{ kg COD/m}^3/\text{day}$. A slight increase in net energy production could be realized by going to a lower loading rate (2 percent more at $6 \text{ kg COD/m}^3/\text{day}$), but the increase in capital expense more than offsets the small gain in energy production.

SUMMARY AND CONCLUSIONS

The performance of the anaerobic fluidized bed system was found to be affected by all the parameters investigated, organic loading rate, influent substrate concentration and temperature. Rather substantial increases in the organic loading rate for whole whey (6 percent solids), however, had only a slight effect. Almost a three-fold increase in loading (13.6 to $37.8 \text{ kg COD/m}^3/\text{day}$) resulted in less than a 12 percent decrease in COD removal efficiency (83.6 to 72 percent). Temperature also had a minor effect. An 11 degree drop from 35° to 24°C resulted in only a 10 percent

TABLE IV. Energy Balance for Processing Whey From a Large Cheese Plant in the Anaerobic Fluidized Bed

Applied Organic Loading Rate (kg COD/m ³ /day)	12	18	24	30	36
Energy produced (kwh/yr)	15.58x10 ⁶	14.85x10 ⁶	14.30x10 ⁶	13.75x10 ⁶	13.38x10 ⁶
Energy Required (kwh/yr)	0.51x10 ⁶	0.34x10 ⁶	0.26x10 ⁶	0.20x10 ⁶	0.17x10 ⁶
Net Energy Produced (kwh/yr)	15.07x10 ⁶	14.51x10 ⁶	14.04x10 ⁶	13.55x10 ⁶	13.21x10 ⁶
Energy Yield Ration ($\frac{\text{Energy Produced}}{\text{Energy Utilized}}$)	30.5	43.7	55.0	68.8	78.7

reduction in COD removal rates over the entire loading range examined (15.0 to 36.8 kg COD/m³/day).

The effect of organic loading rate was most pronounced for an influent substrate concentration of one percent solids. This is to be expected as in addition to diffusional resistances, kinetic limitations begin to reduce the overall reaction rate. The COD removal efficiency was found to vary from 89.5 to 53.3 percent over the range of loadings examined (4.0 to 32.4 kg COD/m³/day).

Based on this information it would appear that reactors in series would prove to be more effective in reducing COD than a single stage system at comparable loadings. This, in fact, was observed when near the end of the study the reactors in series concept was tested. Two reactors loaded at an overall rate of 10.5 kg COD/m³/day provided 94 percent COD reduction (96.5 percent BOD₅ removal). This is 8 to 9 percent greater efficiency than would be anticipated for a single stage reactor at a comparable loading rate. The drawback of the series approach lies in the fact that the relatively high volatile acids levels that would be encountered in the first stage would require greater alkalinity addition to maintain a favorable pH for methanogenesis. Thus in the design of a system a tradeoff of capital versus operating expenses needs to be made.

The fluidized bed system operated in the anaerobic mode can be considered a completely mixed stirred tank reactor (CSTR) (11,12). Due to this the effect of slugs and toxic or inhibitor shocks are mitigated. This also facilitates the control of reactor pH adding to the stability and reliability of the process. The excellent mass transfer properties of fluidized bed, due in part to minimization of the liquid film layer (11-13), enables the system to maintain relative high removal rates despite its completely mixed nature.

Generation of methane gas from cheese whey in an anaerobic fluidized bed system is many times over a net energy producer. Between 30.5 and 78.7 times the energy required to operate the process was produced depending on the applied loading rate. The process therefore seems ideally suited for energy recovery in the cheese industry. Some consideration must be given to ultimate alternate disposal of the system effluent, but it should be kept in mind that this effluent is no higher and in most cases considerably lower in strength than the bottoms stillage generated from alcohol production.

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TREATMENT OF PHENOL WITH AN INNOVATIVE FLUIDIZED
BED ACTIVATED CARBON ANAEROBIC FILTER

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INTRODUCTION

Phenol represents the major constituents in the wastewater effluent of coal gasification and coking plants (1,2). Other sources of phenol-bearing wastewaters include the effluents from smelting and slag process; petrochemical, synthetic resin, pharmaceutical, plywood and fertilizer manufacturing; as well as paint stripping operation (3). The concentrations of phenol present in coal gasification wastewater have been reported to vary from 200-6600 mg/l (1).

Various physical-chemical and aerobic biological processes have been used to successfully treat low concentrations of phenol-bearing wastewaters (3-13). However, the costs involved in treating wastewater containing high concentrations of phenol with these processes could be prohibitively high. This is especially true with the use of physical-chemical processes. The aerobic treatment of phenol-bearing wastewater has frequently been associated with process instability induced by changes in concentration and/or composition of the wastewater (13). In addition, the aerobic treatment requires vigorous bulk liquid mixing and transfer of oxygen to the wastewater, and is known to create

copious amounts of sludge; both aspects often are energy intensive and expensive. By comparison, anaerobic processes not only reduce both of the major operating expenses of an aerobic system, but also, in some instances, especially with the high strength wastewater, result in a net energy gain through methane generation and reduced sludge volume.

The feasibility of using anaerobic system for the treatment of phenol as an alternative to conventional aerobic biological system has been reported by Hobson *et al.* (14). The biokinetics of anaerobic degradation of phenol has been studied by Neufeld *et al.* (15). Chemielowski *et al.* (16) have performed specific kinetic research in the anaerobic decomposition of phenol. Healy and Young (17) have demonstrated the degradation of both phenol and catechol by a methanogenic population of bacteria. The anaerobic biodegradability of phenol and catechol by methanogens has also been confirmed by Khan *et al.* (18) and Suidan *et al.* (19), respectively. However, most of the biological studies on phenol degradation involved an optimum concentration range of a few hundred parts per million (ppm) of phenol in the feed. The treatment process developed in this study combines the advantages of the energy efficient anaerobic filter process, first developed by Young and McCarty (20) and later modified by Chian and DeWalle (21) with recirculation, and the adsorptive capacity of fluidized activated carbon for the long detention of less readily available organic compounds as originally reported by Khan *et al.* (18) for anaerobic degradation of phenol.

The two-stage anaerobic Raschig ring and granular activated carbon packed bioreactors employed in this study for the treatment of synthetic phenol wastewater was first reported by Chian *et al.* (22) on the anaerobic treatment of firefighting wastewater. The advantages of using these systems are that they were originally designed for a pilot-scale operation. As such the results obtained from this study could be easily used for scale-up purpose with minimum modifications required for the design of a larger system. In addition, the use of a first-stage Raschig ring packed roughing filter enhances the initial acclimation of anaerobic degradation of phenol as this stage provides an ample supply of seed necessary for the successful operation of the second-stage fluidized activated carbon column. However, after the system was well acclimated, the sequence of these two reactors can be reversed so that the fixed bed Raschig ring column could serve as a biological filter for the

removal of suspended solids from the activated carbon.

MATERIALS AND METHODS

Two-Stage Reactors

The two-stage pilot-scale anaerobic filter columns employed in this study consist of two identical Plexiglas columns, each having a height of 183 cm and an internal diameter of 10 cm. Each end of these columns was connected to two 20-cm long inverted conical end pieces for inlet and outlet of liquid. A 13-cm diameter concentric Plexiglas water jacket was installed to maintain a constant temperature ($35^{\circ}\text{C} \pm 0.5^{\circ}\text{C}$). The water jackets were connected in series to a constant temperature water bath (American Instrument Model 4-8600, Silver Spring, MD).

The medium packed in the fixed bed consists of 173-cm deep of Raschig rings (1½" nominal size), whereas that packed in the fluidized bed consists of 125 cm of 10x20 U.S. mesh Filtrasorb 400 granular activated carbon (Calgon Corp., Pittsburgh, PA). A 1/3 h.p. stainless steel centrifugal pump (Teel Pump, Dayton, OH) was used to recirculate the aqueous solution and to fluidize the granular activated carbon in order to minimize gas entrainment. Other functions of recirculation include provisions for dilution of influent as well as buffering capacity. The fluidized bed also had an additional 30-cm long by 15-cm diameter expansion chamber at the top to allow for settling of gas-bound carbons.

The feed system includes an influent reservoir containing phenol substrate, ammonium salt nutrient and phosphate buffer solution which were fed to the first column by a variable flow positive-displacement FMI pump (Fluid Metering Inc., Oyster Bay, NY), and to the second column by gravity flow. Both columns were connected to a gas-liquid separator which was interconnected to two 4-ℓ burets for gas collection. Figure 1 represents a schematic diagram of the two-stage anaerobic filters described above.

The feeding substrate employed in this study was an aqueous solution of phenol having a concentration up to 2000 mg/ℓ. During the initial acclimation phase, glucose was added to serve as a readily available carbon source for bacterial growth. Unlike what was reported by Khan *et al.* (18), no vitamins and trace metals were added to the feed. Ammonium chloride and phosphate buffer at a concentration

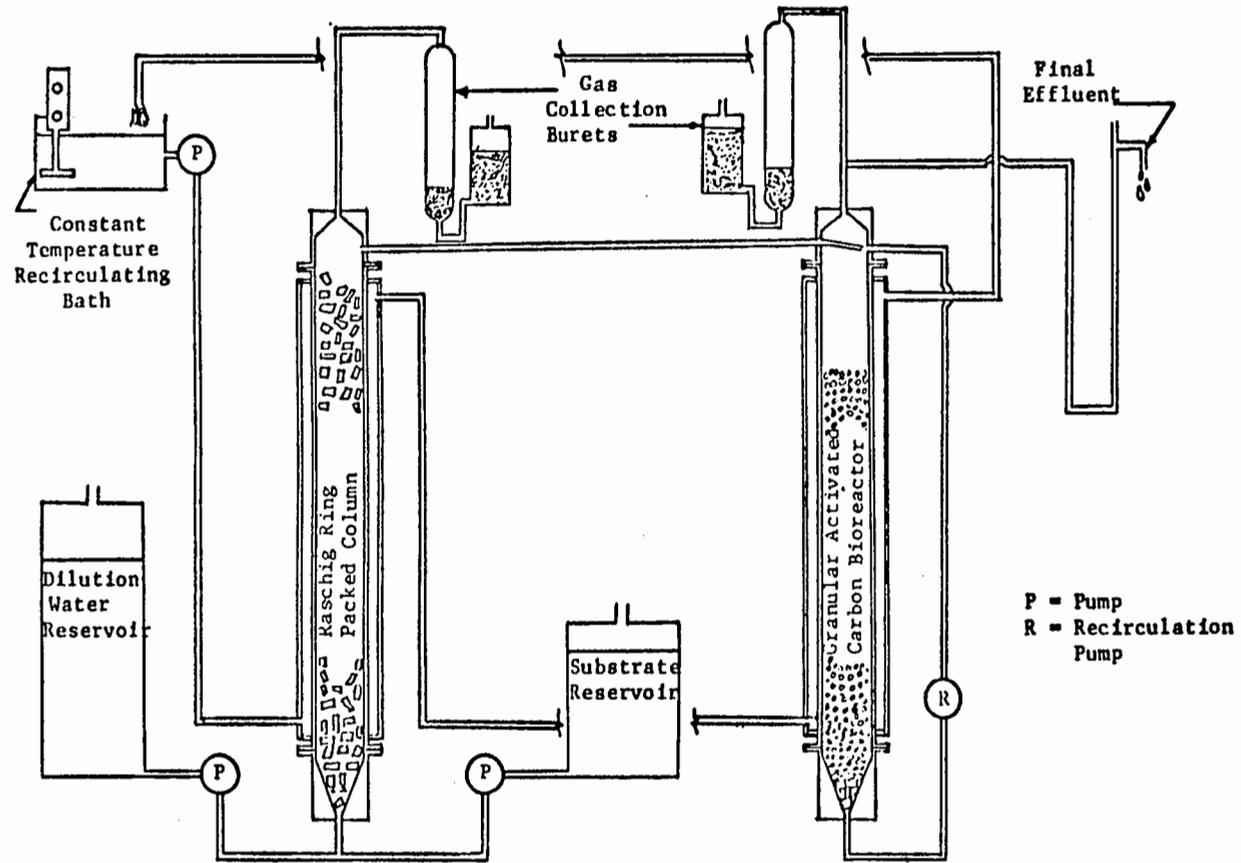


Figure 1. Schematic Diagram of Raschig Ring - Granular Activated Carbon Packed Two-Stage Anaerobic Bioreactor

of 220 mg-N/liter and 335 mg-P/liter, respectively, were the only two other major nutrients added to the synthetic wastewater.

Process Monitoring

Daily checks of pump flow rates (or percentages of carbon bed expansion), feed reservoir volumes, effluent flow rates, pH and gas production were made. In addition, weekly determinations of effluent TOC, COD, alkalinity, total volatile acids and phenol as well as gas composition were carried out in order to assess the performance of each unit in terms of removal efficiency of organic contaminants and the production and conversion of specific compounds, such as organic acids and phenol.

Analytical Procedures

A Fisher Accumet pH Meter, Model 144 (Pittsburgh, PA) was used for pH measurement and a Beckman Model 915 Total Organic Carbon Analyzer (Fullerton, CA) for TOC and TIC determinations. COD and alkalinity were determined according to "Standard Methods" (23). All samples were filtered through a 0.45 μm Gelman Metrical membrane filters (Ann Arbor, MI) prior to analysis.

The gas composition, i.e., methane, carbon dioxide, hydrogen, nitrogen and oxygen were determined with two Fisher Gas Partitioner Model 25V (Pittsburgh, PA) in conjunction with a Fisher Thermal Stabilizer Model 27 and a Coleman Recorder, Hitachi 165 (Pittsburgh, PA).

The specific volatile fatty acids (e.g., C₂ to C₅ acids) were determined with a packed column Hewlett-Packard (HP) Gas Chromatograph Model 4710A equipped with an FID detector and a HP 3380A Integrator (Avondale, PA). The chromatographic column employed in this analysis was an 1/8 in. (0.317 cm) O.D. and 2-ft (61 cm) Pyrex glass coil packed with an acid-washed Carbopack B (60/80 mesh) saturated with 3.0% Carbowax 20M and 0.5% Phosphoric acid. The oven temperature was programmed from 105°C to 150°C at a rate of 4°C/min. The injection port and the detector temperatures were both 250°C. Nitrogen was used as a carrier gas (40 ml/min) while hydrogen (40 ml/min) and compressed air (300 ml/min) were used for the flame ionization detector. All samples were acidified to pH 2 with H₃PO₄ prior to analysis.

Phenol concentration was determined using both UV-Spectrophotometry and Gas Chromatography. The UV absorption was used for the determination of low levels of phenol (less than 100 mg/ℓ). The Gas Chromatographic conditions employed were identical to those used for total volatile acid analysis.

RESULTS AND DISCUSSION

A pilot-scale two-stage anaerobic activated carbon and Raschig ring packed filter system was used in this study. It was operated at a hydraulic retention time of 24 hrs in each column (i.e., a total of 48 hrs through the two-stage system). The Raschig ring packed fixed bed reactor was operated in a plug-flow mode, whereas the activated carbon packed fluidized bed was operated in a back-mixed (i.e., well-mixed) mode. The latter was fluidized by means of effluent recycle at an upflow rate of 5 gpm/ft² (13.3 m³/m²/hr).

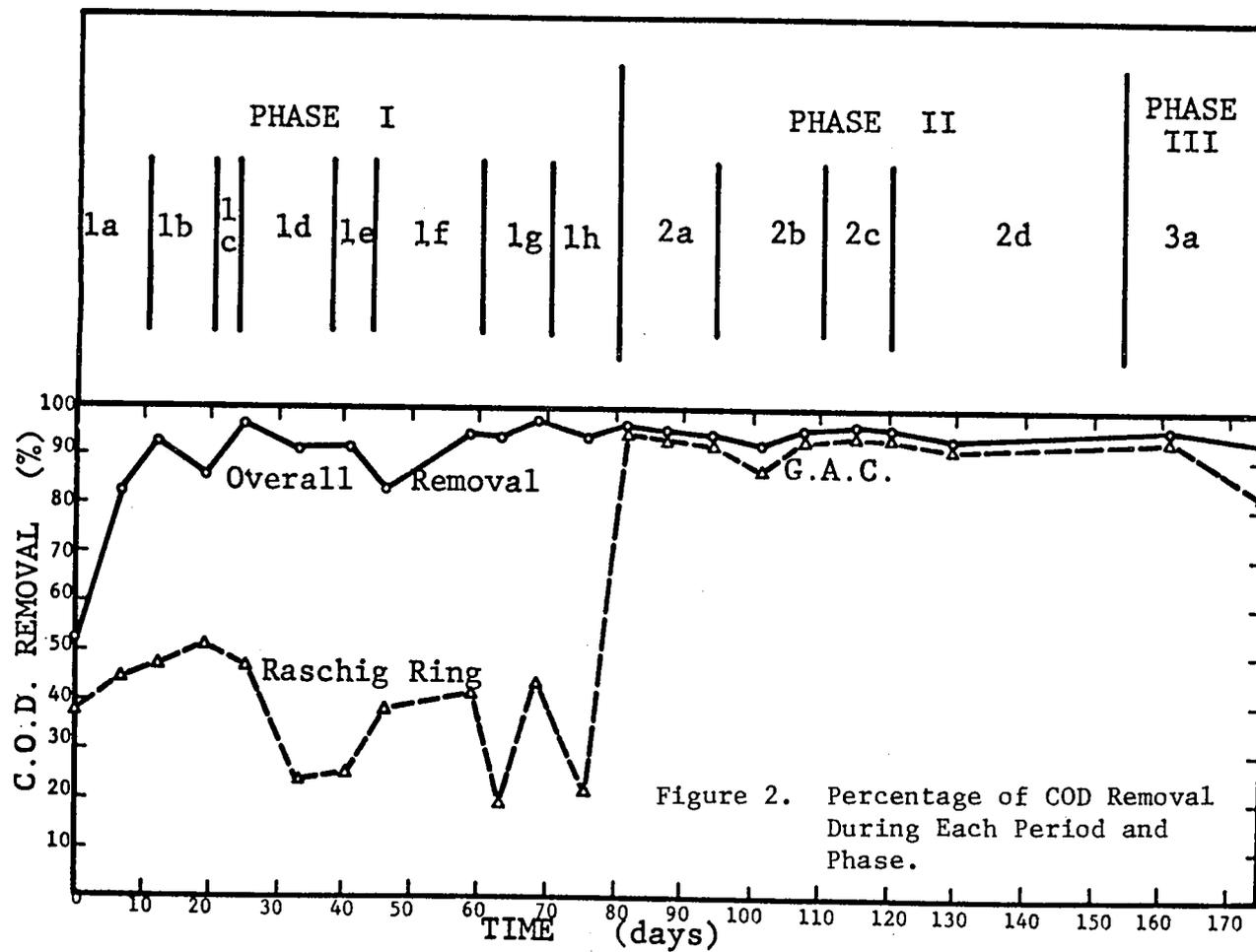
Phase I - Acclimation

During Phase I of the study, which lasted 80 days, the first-stage Raschig ring packed bed was seeded with approximately 4 liters of settled digested sludge (~5-6% solids) collected from a local sewage treatment plant (Clayton Plant, Atlanta, GA). The selection of seeding the Raschig ring packed column was based on previous experience of failure in seeding directly in the fluidized carbon bed. The start-up procedures for acclimating the system receiving phenolic waste are given in Table 1. It is seen from Table 1 that the substrates fed to the system were maintained between 1000-2000 mg/ℓ. During this phase of study, glucose was added to facilitate accumulation of bacterial population capable of producing methane gas. In the meantime, an increasing amount of phenol with a concomitant decrease of glucose concentrations were added to the feed to enhance acclimation of the microorganisms capable of degrading phenol. The use of this start-up procedure was found to greatly accelerate the rate of acclimation of the digested sludge to degrade phenol.

Figure 2 shows the percentage of COD removal during each phase of the study. During the Phase I study, the concentration of glucose was decreased from 800 to 200 mg/ℓ

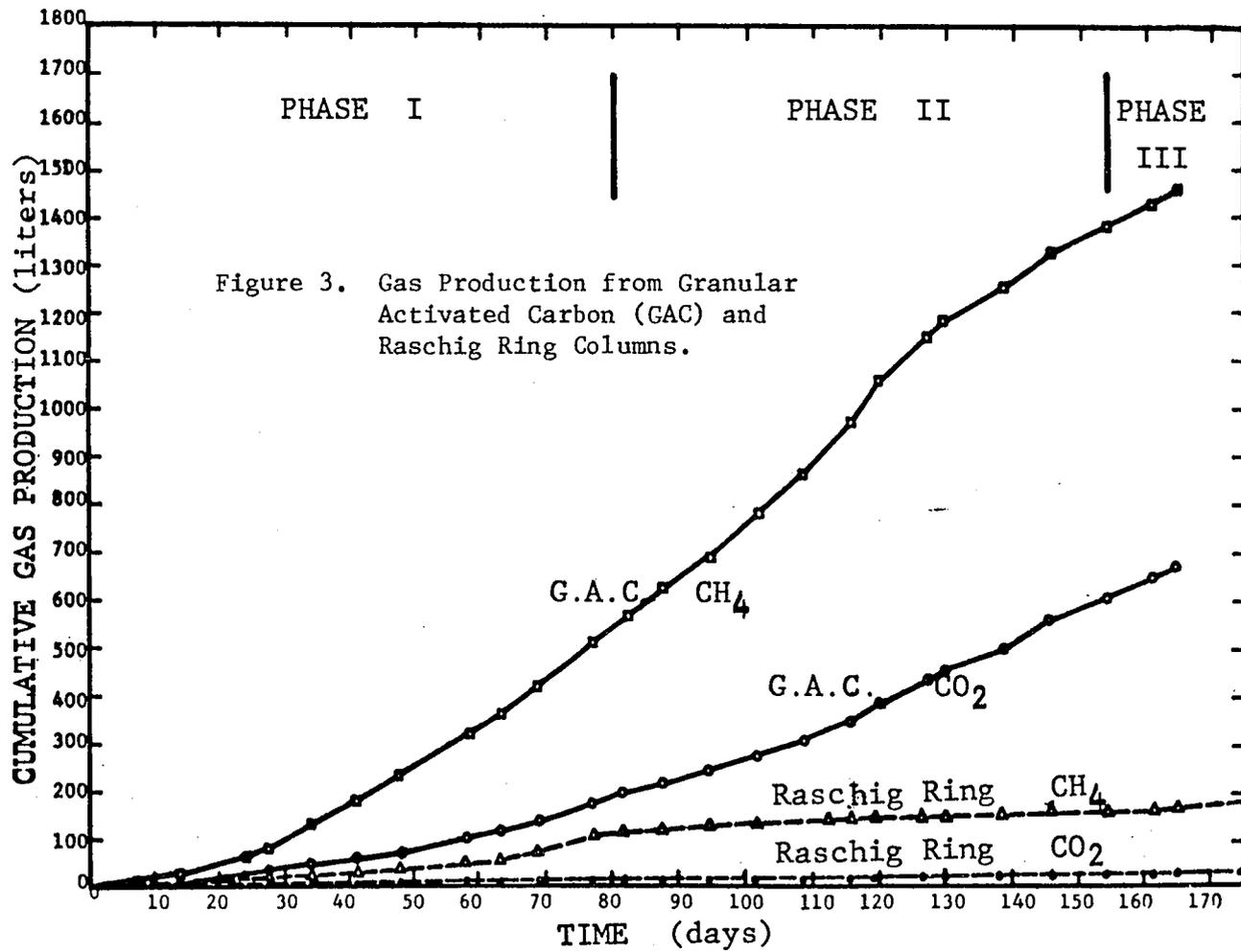
Table 1. Different Phases and Experimental Conditions.

Phase	Duration (days)	Period	Time, day Duration (days)	Feed Concentration (mg/l)			Stage Sequence
				Phenol	Glucose	TOC	
I	80	1a	1-10 (10)	200	800	473	1st stage: Raschig Ring Fixed Bed 2nd stage: Granular Activated Carbon Fluidized Bed Expansion 20% Hydraulic Detention Time 24 hrs (each column)
		1b	10-20 (10)	300	800	550	
		1c	20-24 (4)	300	700	510	
		1d	24-38 (14)	500	600	623	
		1e	38-44 (6)	600	600	700	
		1f	44-60 (16)	700	500	736	
		1g	60-70 (10)	800	300	733	
		1h	70-80 (10)	1000	200	846	
II	74	2a	80-95 (15)	1200	100	959	Switched Sequence: 1st stage: Granular Activated Carbon Fluidized Bed, 20% bed expansion
		2b	95-110 (15)	1400	0	1072	
		2c	110-120 (10)	1600	0	1226	
		2d	120-154 (34)	1800	0	1379	
III	20	3a	154-174 (20)	2000	0	1532	2nd stage: Raschig Ring Fixed Bed



whereas that of phenol was increased from 200 to 1000 mg/l (see Table 1). The rate of substrate variations during periods (Ia) through (Ih) is given in Table 1, and its effect on COD removal efficiencies is shown in Figure 2. The rate of increasing phenol concentrations was based on the pseudo-steady-state production of gas (see Figure 3). As can be seen in Figure 3, the rate of cumulative production of gas, mainly methane and carbon dioxide, increases with the increase in phenol concentrations. During the Phase I acclimation period, this increase is contributed by the overall increase in TOC fed to the system as the theoretical value of TOC to phenol ratio is 0.766 whereas that to glucose is 0.4 (see Table 1). At the end of Phase I study (i.e., day 80, period Ih), the overall removal of COD was 95 percent (Figure 2), whereas that of TOC was 94 percent (Figure 4). However, the overall removal of phenol is near completion (i.e., 99.6%, Figure 5).

It is seen from Figure 2 and 4 that the removal efficiencies of COD and TOC are comparable. An average of 35 percent removal of both TOC and COD was accomplished by the first-stage Raschig ring packed column during Phase I operation. An additional 60 percent removal of both TOC and COD was contributed by the fluidized granular activated carbon (GAC) column. This gives an overall removal of 95 percent of TOC and COD. In the meantime, the overall removal of phenol with the two-stage system was close to 100 percent (see Figure 5). However, the removal of phenol with the Raschig ring column averaged only about 15 percent and reached 25 percent toward the end of the acclimation period (i.e., day 80, period Ih). The complete removal of phenol with the two-stage system even in the early stage of the acclimation period (Phase I) could be the result of the combined effect of adsorption and biodegradation. This is evidenced by a material balance made in the beginning of Phase II study. At a phenol feed concentration of 1200 mg/l, approximately 70 percent of the carbon was accountable from the gas produced and the effluent discharged. This includes both dissolved methane gas and carbon dioxide, and the residual organic compounds present in the effluent. The remainder of 30 percent of carbon is believed to be associated with the biomass accumulated in the system and lost in the effluent (all TOC and COD analyses were done with filtered samples), as well as substrates adsorbed by the activated carbon column. Since the gas phase methane concentration was 68%, the methane yield was calculated to be 48%. By



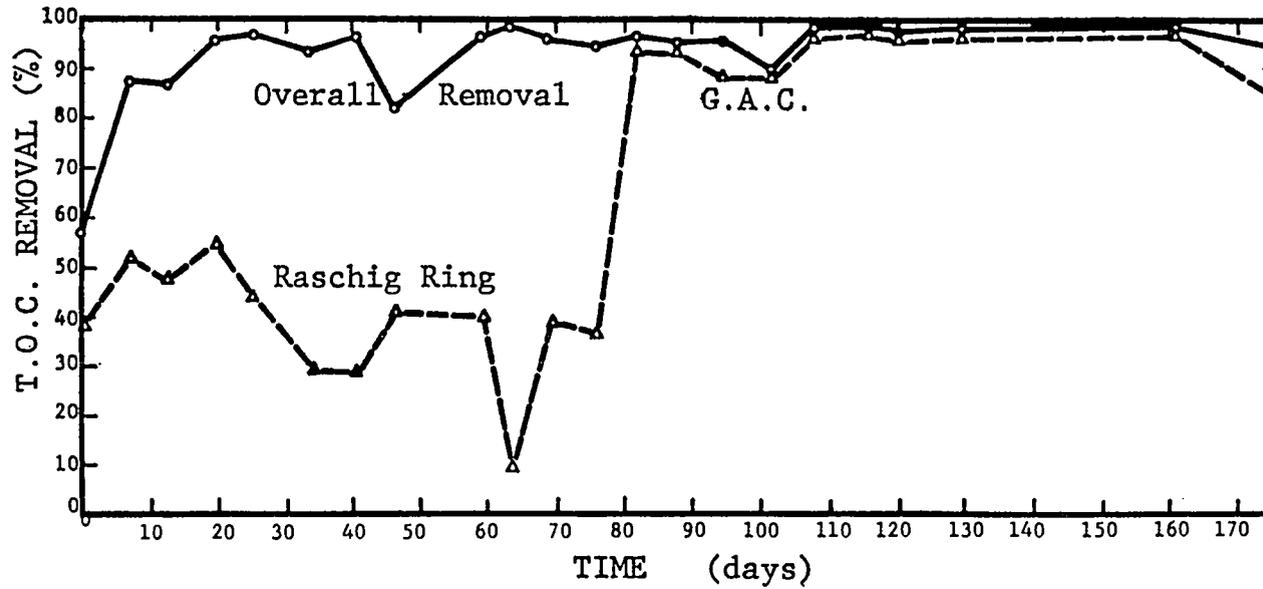


Figure 4. Percentage of TOC Removal During Each Phase.

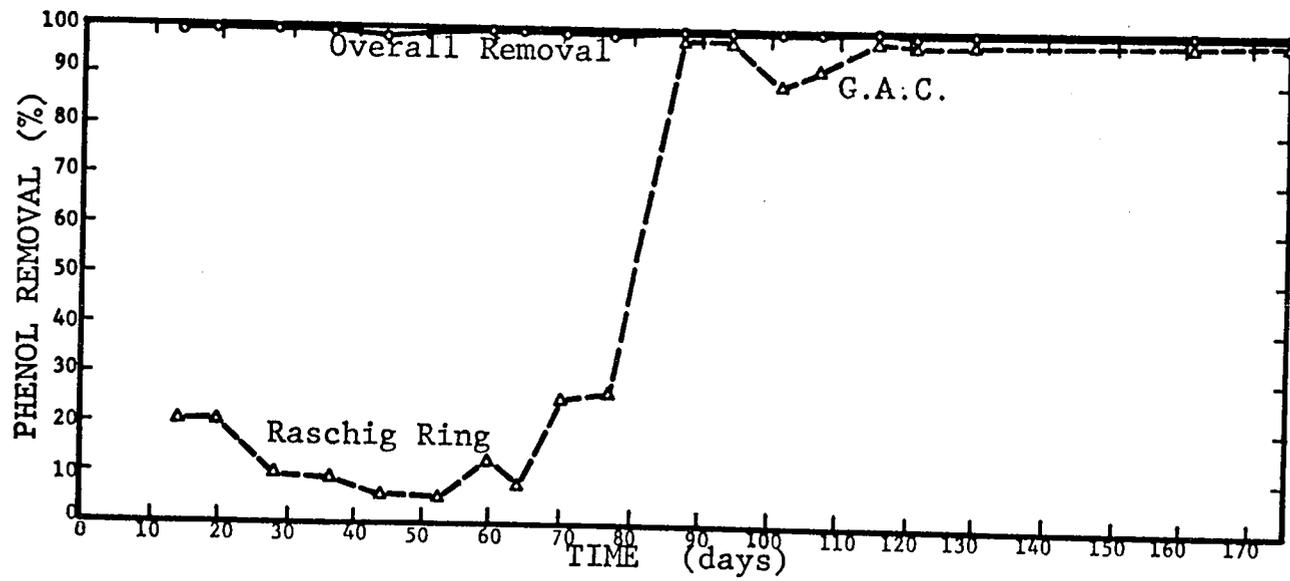


Figure 5. Percentage of Phenol Removal During Each Phase.

considering the solubility of methane gas in aqueous solution, the overall yield of methane was approximately 50%. This is close to the theoretical yield of methane with the anaerobic process.

Phase II Sequence of Reactors

At the beginning of Phase II study (2a, Figure 2), the sequence of the reactor arrangement was reversed, i.e., the feed was first introduced into the fluidized activated carbon column followed by fixed Raschig ring column. During the 15-day period (2a, Table 1), an average of 92 percent of COD removal was actually contributed by the now first-stage granular activated carbon (GAC) column. The removal of COD by passing through the now second-stage Raschig ring column was only 2-3 percent. However, the second-stage fixed bed in turn served well as a biological filter as the effluent suspended solids decreased from 150-200 mg/l range to a somewhat more steady value of approximately 50 mg/l. Since the fixed Raschig ring bed was found in an early study essential in providing a bacterial seed for the fluidized activated carbon column during the start-up period, the role of this stage was therefore to first serve the purpose for acclimation of biomass and then function as a biological filter for suspended solids removal.

During this phase of study, glucose was eliminated totally from the feed (Period 2b, Table 1). A continuing high rate of gas production (Figure 3) was observed when the phenol concentration was increased to 1400 mg/l (Table 1). However, the removal efficiencies of COD, TOC and phenol started decreasing slightly and then returned to the values obtained during the previous period 2a (Figure 2, 4 and 5). This appeared to show the first sign of system upset upon increasing phenol concentration in the feed. At a feed concentration of 1600 mg/l of phenol (Period 2c, Table 1), no sign of any upset of the system was observed in terms of both gas production (Figure 3) and removal of other parameters (figure 2, 4 and 5). However, when the phenol concentration was increased to 1800 mg/l (Period 2d, Table 1), the rate of gas production first showed signs of decreasing (Day 120, Figure 3) although no apparent decrease in the removal efficiencies of other parameters was observed (Figure 2, 4 and 5).

Phase III - Shock Loading

The daily gas production rates were decreased from 25 ℓ /day (day 120) to 14 ℓ /day at the end of period 2d (day 154) when 1800 mg/ ℓ of phenol was fed to the system. A further increase in phenol concentration to 2000 mg/ ℓ was attempted to observe the effect of shock loading to the pseudo stable system. A few days after feeding with a 2000 mg/ ℓ of phenol solution, a significant decrease in the removal efficiencies of both COD and TOC was observed (Figure 2 and 4) with the fluidized activated carbon (GAC) column. A similar decrease in overall COD and TOC removal was also observed although in somewhat lessened extent. The gas production rate dropped to a low of 11 ℓ /day then fluctuating between 11-17 ℓ /day. The overall removal of phenol was, however, maintained unchanged at a maximum of 99.7 percent (Figure 5). This appears that shock loading of phenol could easily be adsorbed by activated carbon. A slight increase in total volatile acids, i.e., from an undetectable level to approximately 50 mg/ ℓ was, however, observed. The accumulation of intermediates reflects the excretion of intermediate metabolites upon shock loading (24,25).

The maximum concentration of phenol evaluated under Phase III study was 2000 mg/ ℓ . Since the system was fed at a rate of 15 ℓ /day and the volume of the fluidized carbon column was 15 ℓ , the column loading was calculated to be 2Kg phenol/ $m^3 \cdot \text{day}$ which is equivalent to 4.76Kg COD/ $m^3 \cdot \text{day}$.

The maximum concentration of phenol reported in the literature was 1000 mg/ ℓ (18) while utilizing anaerobic systems with three fluidized activated carbon columns connected in series. The highest COD loading studied with the innovative two-stage anaerobic filter system employed in the present study was 4.6Kg COD/ $m^3 \cdot \text{day}$ as reported by Chian et al. (22) while treating a firefighting wastewater. However, the highest COD volumetric loading rate reported for the anaerobic degradation of phenol was 9.33Kg/ $m^3 \cdot \text{day}$ (18). This suggests that increased loading should be studied at an optimum feed concentration of 1600 mg/ ℓ phenol. The optimal feed concentration of phenol observed in this study is 4 times higher than the reported value of 400 mg/ ℓ (18).

The decreased percentage of removal of COD and TOC when fed with a high concentration of phenol (2000 mg/ ℓ) as observed in this study was also noted by Chian and DeWalle

(21) while treating a high strength acidic landfill leachate. Contrary to the results in the present study, Jennett and Dennis (26) found that at a constant loading (i.e., 3.5 Kg COD/m³·day), the percentage removal of COD decreased from 98% to 95% and 94%, respectively, when the influent concentrations of a pharmaceutical wastewater decreased from 16,000 mg/l, 8000 mg/l and 4000 mg/l. However, Young and McCarty (20) reached the similar conclusion as observed in this study that the percentage of COD removal decreased as the influent COD concentration increased.

Based on the amount of activated carbon present in the system, the total amount of phenol fed to the system exceeded ten times of the adsorptive capacity of carbon. Biological regeneration of powdered activated carbon has been reported quite frequently with the aerobic biological systems (27). Results of this study also indicate biological regeneration of carbon under anaerobic conditions which confirms with the findings of Khan et al. (18).

CONCLUSIONS

The results presented in the previous sections demonstrated the effectiveness of the two-stage anaerobic filters in the treatment of phenolic wastewaters. Specifically, the following conclusions can be drawn:

1. Results of this study indicated that the anaerobic process employed in this study can be used for the treatment of phenol-bearing wastewater.
2. Whereas the maximum concentration of phenol that can be treated with the anaerobic process is still under evaluation, the optimum concentration of phenol was found to be 1600 mg/l. This is 4 times higher than that reported in the literature.
3. A maximum loading of 2 Kg phenol/m³·day (i.e., 4.76 Kg COD/m³·day) was accomplished with a 94 percent conversion of both COD and TOC.
4. At the maximum loading studied (2 Kg phenol/m³·day) a 99.7 percent conversion of phenol was obtained with the reactors connected in reverse order, i.e., fluidized carbon column followed by fixed Raschig ring bed.
5. By reversing the sequence of reactor order, i.e.,

fluidized column followed by fixed bed, it resulted in an effluent having low suspended solids concentrations.

6. At the optimum phenol concentration of 1600 mg/l (1.6 Kg phenol/m³, day), 97 percent conversion of both COD and TOC and 100 percent conversion of phenol were accomplished.
7. A material balance on carbon indicated that, at a phenol feed concentration of 1200 mg/l, 70 percent of the carbon was accountable from the gas produced and the effluent discharged. The remainder of 30 percent of carbon is believed to be associated with the biomass accumulated in the systems and present in the effluent as suspended solids, and substrate adsorbed by the activated carbon.
8. The first-stage Raschig ring packed bed was found essential in providing a bacterial seed for the second-stage fluidized carbon column during start-up period. After the system was fully acclimated, the reverse order of reactor sequence was desirable in that an effluent of lower effluent phenol concentration and suspended solids resulted.
9. The methane concentration in the gas phase varied from 70 to 85 percent, which could be utilized as a high BTU gas.
10. The observed biological regeneration of activated carbon in the anaerobic system would tend to extend to the useful life of the system indefinitely.
11. It was anticipated that the effectiveness of phenol removal may be impaired by the anaerobic systems fed with an actual wastewater, e.g., coal gasification wastewater, due to the presence of other potentially antagonistic compounds, such as the sulfur and nitrogenous compounds.

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Anaerobic Treatment of Landfill Leachate
By An Upflow Two-Stage Biological Filter

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INTRODUCTION

The sanitary landfill method for the ultimate disposal of solids waste material continues to be widely accepted and used. Selection of the site must be given special attention so that the landfill is properly designed, constructed, and operated. After operation of the fill has started, problems may still develop. One such problem is caused by rainwater which infiltrates into the landfill and subsequent movement of liquid or leachate out of the fill into surrounding soil. This leachate is hard to categorize as it can have a wide range of concentration of chemically diverse contaminants. The leachate however often contains a high concentration of organic matter and inorganic ions, including heavy metals (1).

Pollutions of groundwater and municipal water supply by leachate have been reported during the past (2, 3). The only way to avoid and correct these situations is to collect and treat the leachate. Collection of leachate is a relatively simple engineering task. The problem comes when one tries to treat the leachate properly and economically.

According to Uloth and Mavinic, high-strength leachate wastewater could be effectively stabilized by activated sludge operated at the organic loading rate = 0.8 to 4.8 kg COD/m³-day. The range of COD and BOD removal was within 96.8- 99.2% (4). Chian and DeWalle also studied the activated sludge treatment for lysimeter-generated leachate at organic loading ranged from 1.15 to 5.02 kg COD/m³-day, and found that at least 97% of the influent COD was removed within a mean cell residence time as short as 7 days (5). Recently Zapf-Gilji and Mavinic have reported that the activated sludge can effectively remove not only organic matter but also metal ions contained in leachate (6).

Treatment of sanitary landfill leachate by anaerobic filter system was earlier studied by many investigators. Ham and Boyle found that anaerobic filter system could stabilize a raw leachate of approximately 10,000 mg/l as COD with a retention time of 10 days and loading less than 32 lb COD/1000 ft³-day (0.512 kg/m³-day). Their study indicated a 90% COD reduction. As the system retention time and the loading were increased and decreased to 12.5 days and 13 lb COD/1000 ft³-day (0.207 kg/m³-day), respectively, the COD removal was increased to 93% (7). In addition, Foree and Reid obtained a COD reduction of 96% at the loading rate equal to 80.2 lb COD/1000 ft³-day (1.283 kg/m³-day)(8). Chian and DeWalle further stated that a high-strength leachate with an acidic pH could be treated using a completely-mixed anaerobic filter with flow recirculation (9). In general, the advantages of anaerobic treatment include: (a) less sensitive to shock loading, (2) high treatment efficiency and more stable effluent, (3) low solids production, and (4) energy production.

This work was designed to study the feasibility of using a two stages upflow anaerobic fixed-film reactor for the treatment of acidic leachate wastewater. The effects of organic loading, reactor retention time, and metal ions on the removal of BOD, COD and the production of methane gas were studied.

EXPERIMENTAL PROCEDURES

A laboratory scale two-stage anaerobic submerged biological filter (AnSBF) was constructed in the Environmental Engineering Research Lab at the University of Pittsburgh. A schematic flow diagram of the pilot plant is shown in Figure 1.

An ultraMaster pump with speed control was employed for feeding the leachate wastewater to the first stage of the anaerobic filter system. The leachate flew up through and out the filter where samples were taken and the gas was measured.

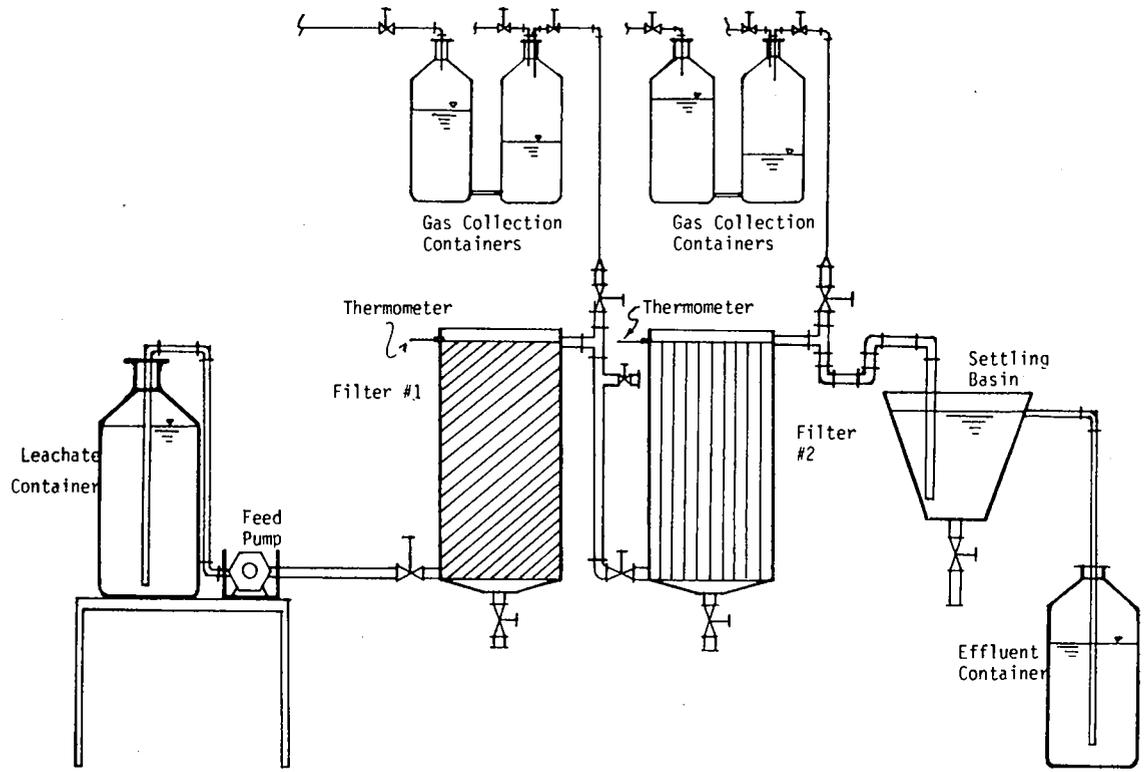


Figure 1. Schematic Diagram of the Laboratory Fixed-Film Anaerobic Filter System

Then, the leachate continuously ran through the second filter by the action of the feed pump. After passing through the second filter, the treated leachate was collected in a settling basin where both substrate and solids concentrations were measured. Gas samples were also taken from the second filter.

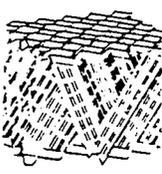
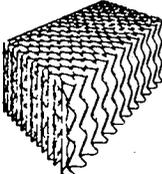
The detailed information regarding the size of the anaerobic submerged biological filter (AnSBF) and the structure of the filter media were reported in Table 1. Influent and effluent samples were taken three times a week. Parameters tested included BOD, COD, total and volatile suspended solids (TSS & VSS), alkalinity, volatile acid as acetic, pH and ammonia nitrogen ($\text{NH}_3\text{-N}$). All of these measurements were conducted in accordance with "Standard Methods" besides pH and ammonia nitrogen (10). These two parameters were determined by using an Orion Digital Ionalyzer. The metal ions, Fe(III) and Zn (II) were measured by an Atomic Absorption Spectrophotometer. Gas Production was determined in two 20-liter containers connected at the bottom and filled with water. The first container was directly connected to the gas line while the second container was open to the atmosphere. The amount of displaced water was the volume of gas produced. The percentage of methane in the gas was measured by a combustible gas indicator. The procedure employed for methane gas determination was also indicated in "Standard Methods".

RESULTS AND DISCUSSION

(A). Leachate Characteristics. The leachate wastewater used in this study was obtained from Elizabeth Township Municipal Landfill located along the Monogahela River 20 miles south of Pittsburgh, Pennsylvania. This landfill is 16 years old and the leachate content is comparatively weak in nitrogen, phosphorus, alkalinity, and metal ions, but strong in volatile acid and solids. Chemical analyses of raw leachate found that the concentrations of nitrogen, phosphorus, alkalinity, metal ions, acetic acid, and suspended solids were approximately equal to 100 mg/l as $\text{NH}_3\text{-N}$, 20 mg/l as $\text{PO}_4\text{-P}$, 1,000 mg/l as CaCO_3 , 20 mg/l as Fe(III), 20 mg/l as Zn(II), 2,480 mg/l as acetic, and 2,100 mg/l as TSS. The leachate wastewater is acidic and it has a pH nearly equal to 5.0.

Due to the deficiency of nitrogen and phosphorus in the raw leachate, ammonia chloride and monobasic potassium phosphate were added to the fresh leachate solution. The amount of nitrogen and phosphorus supplemented was kept to maintain a ratio of COD:N:P between 100:1:0.1 and 100:2:0.2 in the feed solution. Previous work of Chian and DeWalle reported that the COD:N:P ratio of 100:2.7:0.2 was adequate for successfully

Table 1.. Anaerobic Filter Plant Specifications

Parameters	Filter #1	Filter #2
(A). Reactor Dimension:		
Size -----	1.0' x 1.0'	1.0' x 1.0'
Height -----	2.0'	2.0'
Volume		
Gross -----	2 ft ³ (56.6 l)	2 ft ³ (56.6 l)
Net -----	1 ft ³ (28.3 l)	1 ft ³ (28.3 l)
(B). Filter Media:	Munters Bjo- dek 19060*	B. F. Goodrich Koro-z**
Surface Area -----	42 ft ² /ft ³	44 ft ² /ft ³
Approx. Void Vol. ----	97.2%	96.0%
(C). Media Configuration:		

* Flow is constantly redistributed in the horizontal direction.

** Flow is constantly redistributed in the vertical direction.

treating the landfill leachate by a submerged anaerobic filter if flow recirculation was employed (9).

Table 2 lists the general characteristics of landfill leachate employed for the present study. And daily influent leachate conditions are shown in Figure 2 thru Figure 4. It can be seen in table 1 that the influent COD and BOD concentrations in the feed were approximately the same for phases I and II of the study. The COD and BOD averaged around 21,000 mg/l and 14,000 mg/l, respectively, while they never varied more than 4,600 mg/l above or below the average. For phase III study the influent COD was reduced to an average of 9,604 mg/l. As a result, the averaged BOD concentration was decreased to 7,227 mg/l. According to table 1, the TSS and VSS were 1,829 mg/l and 1,413 mg/l for phase I, 1,742 mg/l and 1,402 mg/l for phase II, and 1,385 mg/l and 1,209 mg/l for phase III. With the addition of sodium bicarbonate, the alkalinity in leachate feed was 2,287 mg/l, 2,621 mg/l, and 2,208 mg/l for phases I, II, and III, respectively.

Since the AnSBF was built into two stages in series, the effluent of filter 1 is the influent of filter 2. The influent properties of the filter 2 were determined by measuring the physical and chemical compositions of the treated leachate after passing through the filter 1. The data describing the influent conditions of the filter 2 was also shown in table 2.

(B). Operating Conditions. The two stages pilot-scale AnSBF was operated under three different loading rates as shown in table 2. In the filter 1, the organic loadings were 142 lb COD/1000 ft³-day (2.272 kg/m³-day) or 69 lb BOD/1000 ft³-day (1.104 kg BOD/m³-day) for phase I, 75.2 lb COD/1000 ft³-day (1.203 kg COD/m³-day) or 56.4 lb BOD/1000 ft³-day (0.902 kg/m³-day) for phase II, and 62.3 lb COD/1000 ft³-day (0.997 kg/m³-day) or 48.5 lb BOD/1000 ft³-day (0.774 kg BOD/m³-day) for phase III. The reactor retention times for the above three phases were 9.07 days, 16.7 days, and 9.5 days, respectively. All of the organic loading conditions used to operate the filter 2 were also reported in table 2.

Figure 5 shows the daily COD and BOD loading rates. It is clearly indicated in Figure 5 that the organic loadings applied to filter 2 during the first 20 days of phases II and III study fluctuated greatly because of a non-steady state operational condition in filter 1. However, three loading rates used to operate the filter 1 were kept fairly constant throughout each phase of the present study.

(C) System Start-up. Start-up of an anaerobic filter probably is the most difficult period of operation. Start-up time in experimental full-scale units has ranged from 10 to

Table 2. Leachate Characteristics and Anaerobic Filter Operating Conditions

Parameters	Anaerobic Filter No. 1					
	Phase I		Phase II		Phase III	
	Range	Mean	Range	Mean	Range	Mean
Influent Waste-water Property:						
Soluble COD, mg/l	17,775-24,500	21,476	17,545-24,900	20,244	8,350-10,690	9,604
Soluble BOD, mg/l	10,575-16,050	13,496	11,250-18,900	14,374	6,240-8,260	7,227
NH ₃ -N, mg/l	101-278	230.9	150-220	182.8	200-305	271.0
TSS, mg/l	1,010-2,400	1,829	1,060-2,900	1,742	1,170-1,660	1,385
VSS, mg/l	940-2,121	1,413	820-2,800	1,401	840-1,470	1,209
Alkalinity as CaCO ₃ , mg/l	1,010-2,339	2,287	1,661-3,370	2,621	1,668-2,897	2,208
pH	4.2-5.8	5.0	4.2-5.9	4.9	4.3-5.8	5.09
Organic Acid as Acetic Acid, mg/l	1,900-3,520	2,890	1,468-3,502	2,545	1,479-2,718	1,940
(BOD/COD)/(VSS/TSS), %	-	62.8/77.2	-	71.0/80.4	-	75.0/87.2
Filter Operating Conditions:						
COD Loading 1b COD/day/1000 ft ³	105.8-198.0	142	58.6-91.4	75.2	53.2-77.4	62.3
BOD Loading 1b BOD/day/1000 ft ³	68.4-142.0	91.7	43.0-78.2	56.4	40.0-63.6	48.5
Retention time, days	5.2-10.7	9.07	9.9-22.6	16.7	6.3-13.9	9.5
Temperature, °C	35-39	37.1	35.5-39	36.9	36.0-37.5	36.5

Table 2. Continued.

Parameters	Anaerobic Filter No. 2					
	Phase I		Phase II		Phase III	
	Range	Mean	Range	Mean	Range	Mean
Influent Waste-water Property:						
Soluble COD, mg/l	9,450-15,750	9,006.5	804-11,705	3,878.6	700-2,404	1,392.5
Soluble BOD, mg/l	5,650-12,480	7,335.0	500-9,295	2,313.5	177-1,366	633.1
NH ₃ -N, mg/l	154-445	274.7	225-380	313.3	239-425.0	350.0
TSS, mg/l	170-3,036	2,092.0	380-2,480	1,065.0	80-335	206.0
VSS, mg/l	135-2,434	1,264.4	220-1,900	721.1	75-285	129.2
Alkalinity as CaCO ₃ , mg/l	4,856-7,800	5,904.2	6,546-9,858	8,170.4	1,615-9,025	6,980.3
pH	6.1-7.8	6.69	7.6-8.4	8.05	7.45-8.40	7.85
Organic Acid as Acetic Acid, mg/l	4,770-11,300	8,172.8	469-4649	1,454.8	281.2-556	368.2
(BOD/COD)/(VSS/TSS), %	-	81.4/60.4	-	59.9/67.7	-	45.5/62.7
Filter Operating Conditions:						
COD Loading lb COD/day/1000 ft ³	78.4-150.2	106.7	2.8-43.2	13.96	4.2-13.7	7.92
BOD Loading lb BOD/day/1000 ft ³	46.8-103.5	69.0	1.81-33.4	8.33	0.88-7.52	3.25
Retention time, days	5.2-10.7	9.07	9.9-22.6	16.70	6.3-13.9	9.50
Temperature, °C	35.0-41.0	38.30	36.0-39.5	37.40	36.5-38.0	36.80

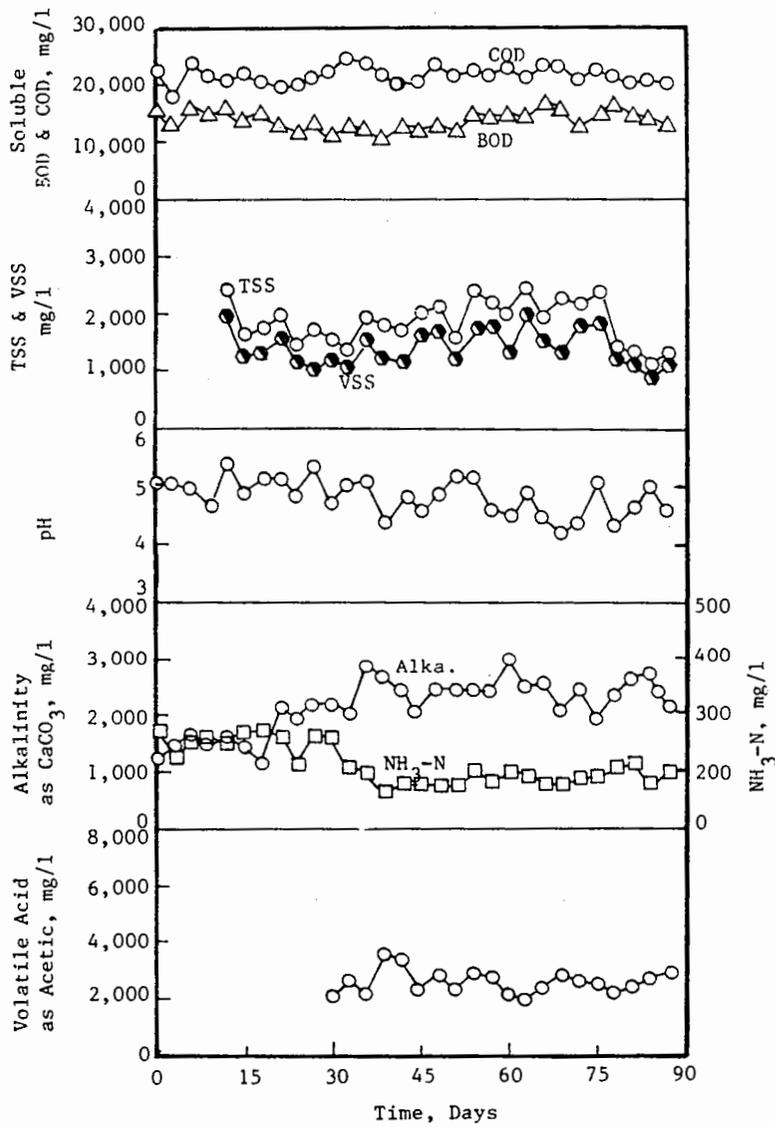


Figure 2. Influent Leachate Composition In Phase I Study

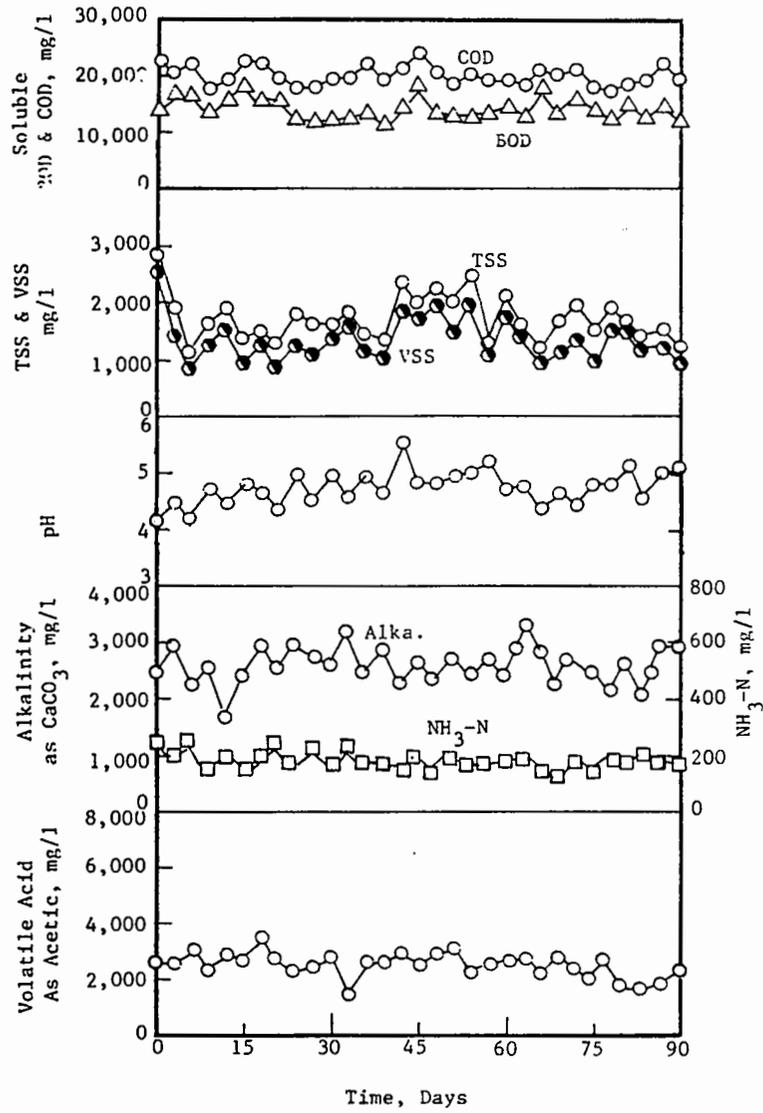


Figure 3. Influent Leachate Composition In Phase II Study

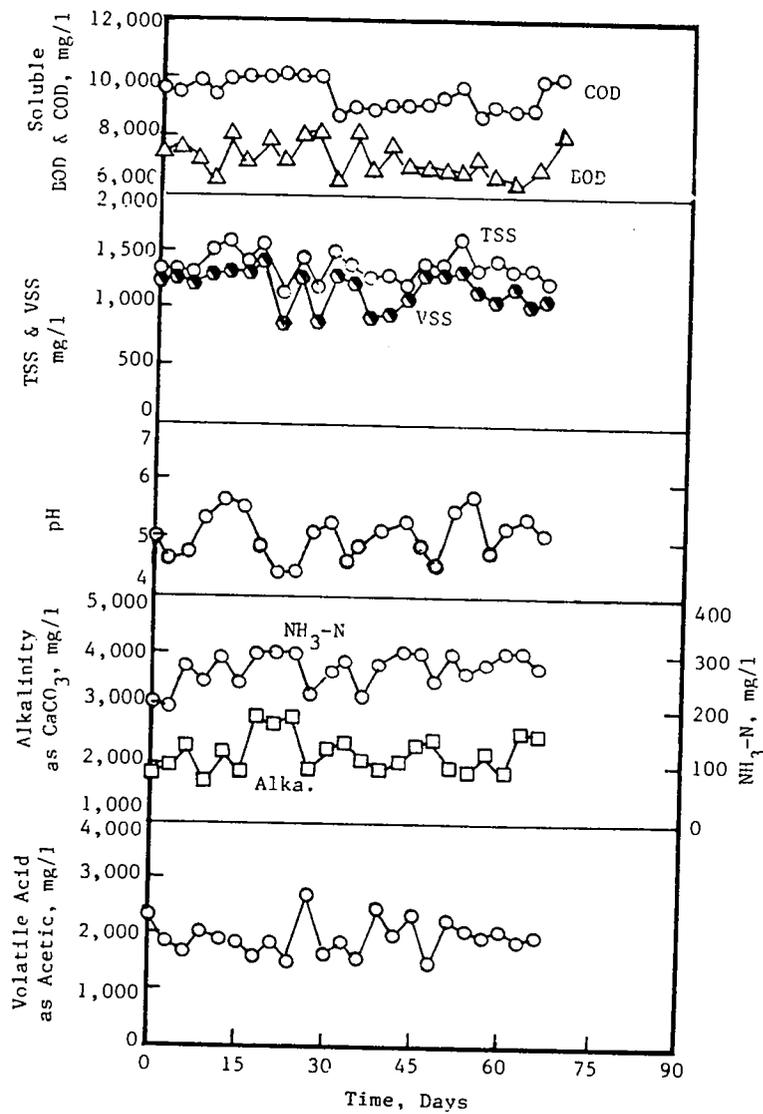


Figure 4. Influent Leachate Composition In Phase III Study

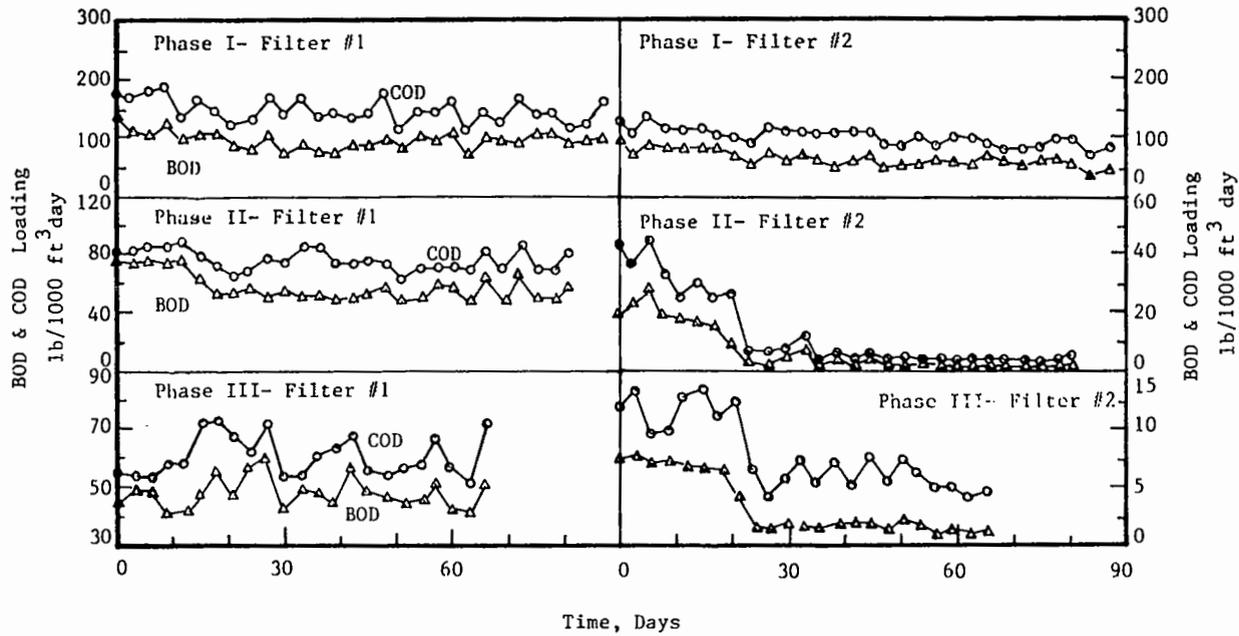


Figure 5. Influent Organic Loadings In Phases I, II, and III Study

180 days with the shorter times corresponding to the use of large amounts of active seed while the longer times were associated with the use of light seeding (11).

20 liters of anaerobic digester sludge was obtained from Pleasant Hill sewage treatment plant at Pleasant Hills, Pennsylvania and 10 liters (80 g. solids) was added to each filter. A synthetic leachate recommended by Stanford, Ham, and Anderson was fed to the filter system for approximately 5 weeks. This leachate was made up of sodium acetate, glycine, acetic acid, pyrogallol, and ferrous sulfate. Very little gas was produced. A new 20 liters seed was obtained and fed into the filter again. Leachate from the Elizabeth Township Municipal Landfill along with chemical additions, was used to feed the system. After approximately seventy days a steady-state condition was first obtained (see Figures 6 and 9). It should be noted that the filter was not heated during the feeding of the synthetic leachate as it was for the leachate collected from the above mentioned landfill site. The start-up time could have been reduced if the filters would have been heated then.

Analysis of the AnSBF plant performance during the start-up period has indicated three factors in importance. First, the slow growth of anaerobic microorganisms, especially at low waste load and at temperature below 30°C, does not permit rapid build-up of biological solids. Consequently, a large seed mass is needed for rapid start-up. Secondly, pH drops below 6.5 at any point within the filter increases the starting time significantly. The results of the present study as seen in Figure 6 indicate that before the steady-state condition reaches, the pH always is below 6.5 in filter 1, due to the volatile acid built up. As the system approaches the steady-state condition, both alkalinity and pH start increasing with decreasing the volatile acid concentration. This means that the system starting time can be reduced if the buffer capacity of the waste is strengthened by the addition of alkalinity. A third factor affecting start-up time is related to the physical characteristics of fixed-film supporting media. During the early stages of operation a significant fraction of the biological solids remain finely dispersed throughout the liquid phase and a considerable fraction washes out with the filter effluent. The structure and configuration of the filter media will determine the rate of cell attachment and development.

However, it is also important to point out the fact that although large seed volume can shorten the starting time, it may contribute significant amounts of inert volatile and non-

volatile solids that tend to plug the filter and reduce its effectiveness for treating wastes. It is necessary to investigate the relationship between seed volume and filter void volume so that the optimum seed condition can be determined.

(D). Substrate Removal. Figure 6 thru Figure 8 show the filter plant effluent conditions for phases I, II, and III study. In phase I the effluent conditions in terms of COD, BOD, TSS, VSS, pH, alkalinity, and volatile acid became stable after seventy days operation. Due to the changes in organic loading condition in phases II and III, the first twenty two days data varied greatly. A transient period occurred when the organic loading rate and/or the reactor retention time changed.

The treatment efficiency of the filter plant was summarized in table 3. All data as shown in table 3 were calculated using the last 30 days results obtained from each phase of the study. This was made because the last 30 days of each phase produced the most consistent data.

(a). COD and BOD removal. From table 3, it can be observed that phases II, with organic loading of 62.3 lb COD/1000 ft³-day or 48.5 lb BOD/1000 ft³-day, obtained the highest substrate removal at 96% as COD and at 98.4% as BOD. Phase I, which had a loading of 142 lb COD/1000 ft³-day or 69 lb BOD/1000 ft³-day only obtained an average COD and BOD reductions of 68.8% and 77.5%, respectively.

Further analysis of COD and BOD data found that filter 1 was more efficient than filter 2 in COD and BOD removal. For instance, in phase II 93% of COD and BOD was removed by filter 1 and only 39.8% of COD or 54.2% of BOD remaining in the effluent of the filter 1 was removed by filter 2. Previous studies reported by others have also shown that the first few feet of filter depth removes the most organic substrate (12, 13). By observing the BOD/COD ratio in table 3, the ratio generally decreased from influent to filter 1 to filter 2 indicating that the leachate became less degradable as it passed through filter 1.

(b). Nitrogen Removal. According to the present study, the AnSBF was not capable of removing nitrogen as was hoped. Ammonia nitrogen build-up was found. This result certainly explains why the nitrogen requirement for leachate treatment by anaerobic filter process is low.

(c). Changes In pH, Alkalinity, And Volatile Acid. The averaged pH of the leachate wastewater for the three phases were 4.73, 4.90, and 5.14. As the wastewater passed through the filters 1 and 2, the pH steadily increased with increasing

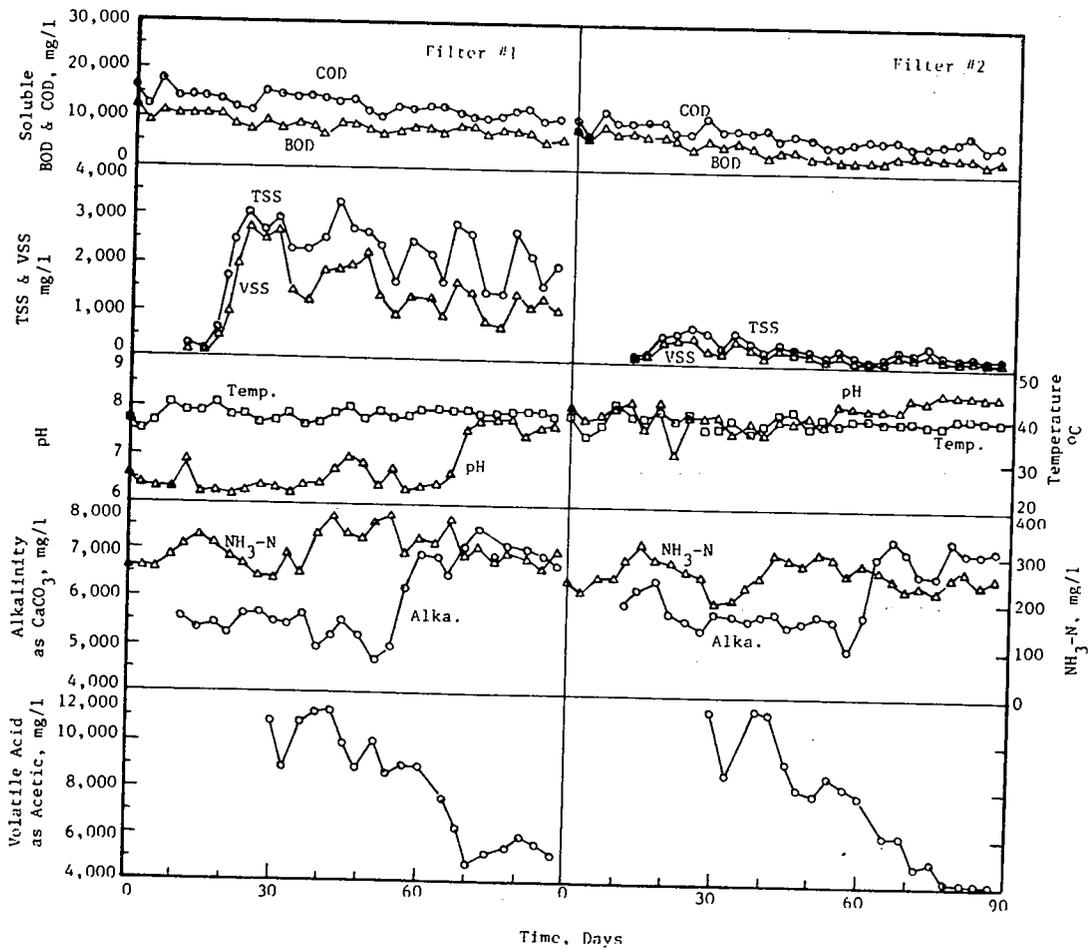


Figure 8. Pilot Plant Effluent Condition In Phase I Study

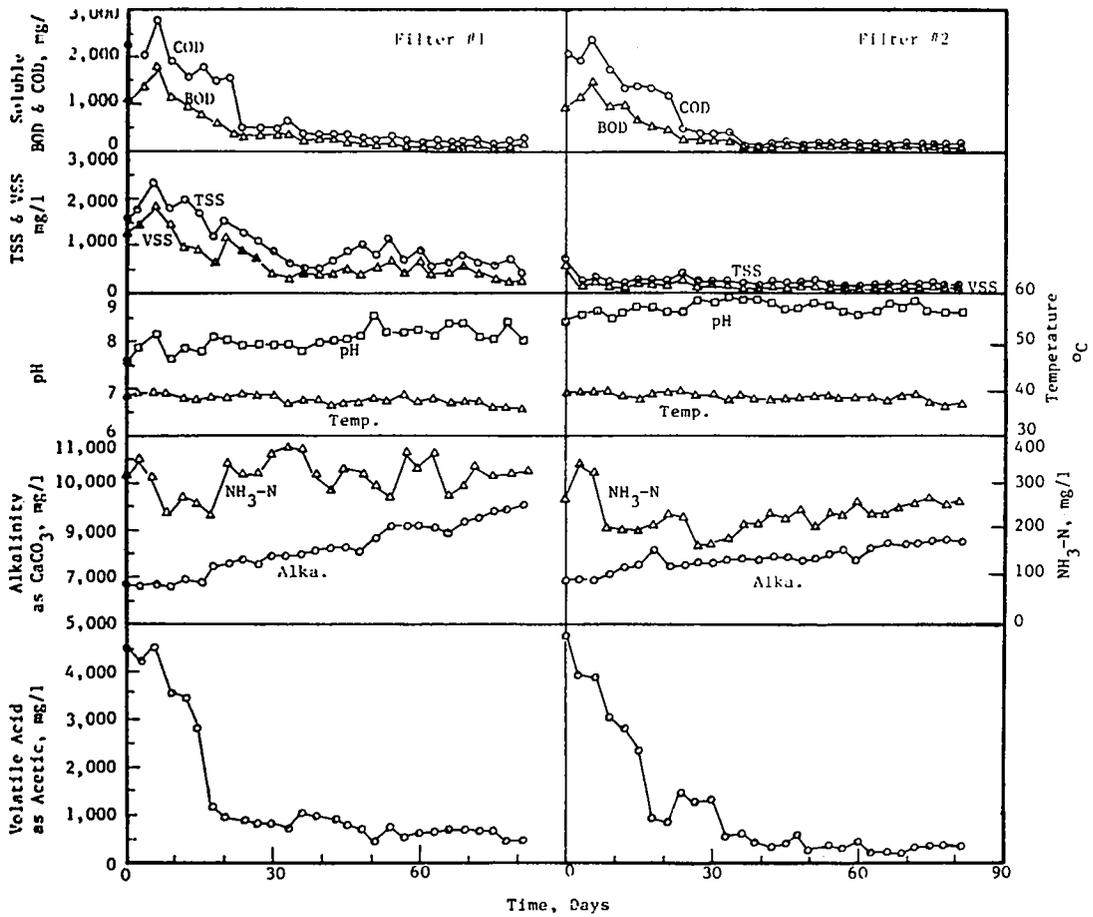


Figure 7. Pilot Plant Effluent Condition In Phase II Study

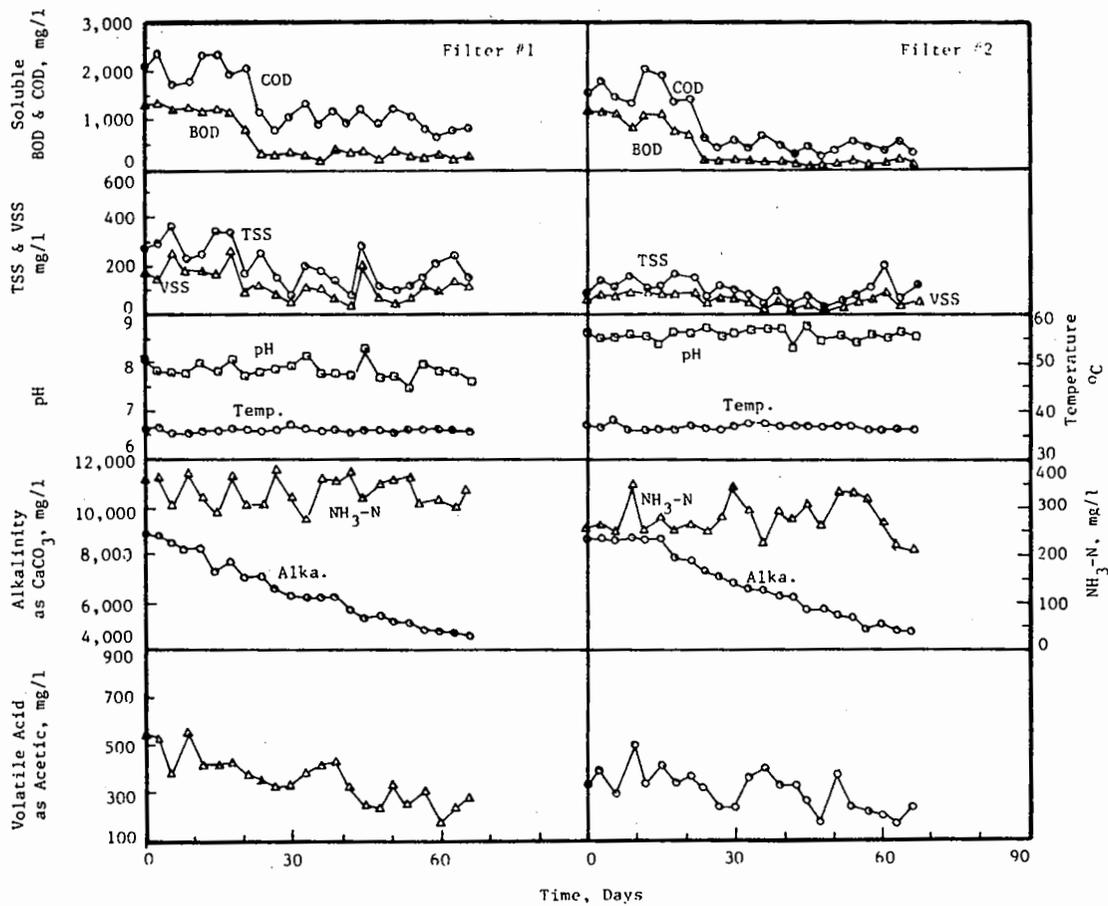


Figure 8. Pilot Plant Effluent Condition In Phase III Study

Table 3. Two-stage Anaerobic Filters Performance*

Parameters	Phase I				Phase II				Phase III			
	Inf.	Filter #1	Filter #2	Overall	Inf.	Filter #1	Filter #2	Overall	Inf.	Filter #1	Filter #2	Overall
Soluble COD, mg/l	21,868	11,320	6,820	15,048	19,571	1,370	824.6	18,746.4	10,117	1,035	401.6	9,715.4
% COD Reduction	-	48	39.7	68.8	-	93	39.8	95.7	-	89.7	61.2	96
Soluble BOD, mg/l	14,043	7,239	3,154	10,889	12,971	908.0	415.6	12,555.4	6,988	244.4	106.5	6,881.5
%BOD Reduction	-	48.2	56.4	77.5	-	93	54.2	96.7	-	96.5	56.4	98.4
NH ₃ -N, mg/l	182.7	314.1	244.1	(-)	187	333.5	219.6	(-)	276.6	333.5	284.0	(-)
% NH ₃ -N Reduction	-	(-)	22.2	(-)	-	(-)	35.2	(-)	-	(-)	14.8	(-)
TSS, mg/l	1,830	2,115.	230.2	1,599.8	1,749	780.6	182.0	1,567.0	1,376	156.0	89.6	1,285.4
% TSS Reduction	-	(-)	90.6	87.4	-	55.3	76.6	89.5	-	88.6	42.5	93.4
VSS, mg/l	1,473	1,561.	164.6	1,608.4	1,380	486.8	112.8	1,267.2	1,165	98.5	33.0	1,132.0
%VSS Reduction	-	(-)	89.5	90.7	-	64.7	76.8	91.8	-	91.5	66.4	97.1
Alkalinity as CaCO ₃ , mg/l	2,404	6,234.	6,334	2,614.0	2,614	8,915	7,476.4	(-)	2,200	5,605	5,706	(-)
% Alkalinity Reduction	-	(-)	(-)	(-)	-	(-)	16.1	(-)	-	(-)	(-)	(-)
Volatle Acid as Acetic Acid, mg/l	2,692	7,195.0	6,168	(-)	2,569	728.5	373.3	2,196.2	1,966	319.1	274.0	1,692.0
% Organic Acid Reduction	-	(-)	14.3	(-)	-	71.6	48.7	85.4	-	83.7	14.7	86.1
pH	4.73	7.03	8.18	-	4.9	8.16	8.73	-	5.14	7.79	8.56	-
BOD/COD, %	64.2	63.9	46.20	-	66.2	66.2	50.4	-	69.0	23.6	28.40	-
VSS/TSS, %	80.5	73.8	71.50	-	78.8	62.3	61.9	-	84.6	63.1	36.80	-

-Note- * Average of last 30-day data, (-) Negative Value

- (1) The values were calculated by using the concentrations of influent and filter #1 effluent.
- (2) The values were calculated by using the concentrations of filter #1 and filter #2 effluent.
- (3) The values were calculated by using the concentrations of influent and filter #2 effluent.

alkalinity and decreasing volatile acid. The pH of the leachate, after it traveled through filter 1, rose to 7.03 in phase I, 8.16 in phase II, and 7.79 in phase III. The final effluent pH, according to table 3, were 8.18, 8.73, and 8.56, respectively, for phases I, II, and III.

As mentioned earlier, the alkalinity of the influent leachate wastewater was kept above 2,000 mg/l. This insured an adequate buffering capacity for radical pH changes. As the leachate stayed in the filter longer, the alkalinity became higher because of pH increase and CO₂ production. The averaged alkalinity in the filter 2 effluent was much higher than that in the influent of filter 1.

Volatile acid as acetic in the influent wastewater averaged 2,691 mg/l, 2,569 mg/l, 1,966 mg/l for phases I, II, and III. As the leachate entered the filter, the COD was biologically broken down into organic acids and they were in turn converted to methane and CO₂. Therefore, there should be a large concentration of volatile acids at the bottom of the filter and this concentration should decrease as the residence time of the leachate wastewater increased. The percent acetic acid reduced in phases II and III was 85.4 and 86.1. However, the acetic acid concentration shows a net increase in phase I which could explain why the COD reduction was the lowest among three phases. Clearly the system was overloaded in this case.

(D). Solids Removal and Sludge Production. Although the contents of TSS and VSS were high in raw leachate, the AnSBF was able to reduce more than 87% of both types of solids. In general, when the solids content was high in the effluent of filter 1, filter 2 becomes very effective in the removal of suspended and volatile solids. Otherwise, filter 1 always produced a higher % solids removal than filter 2.

The ratio of VSS to TSS in the untreated leachate was nearly to 0.813 or 81.3%. As expected, the ratio should decrease after anaerobic treatment. Another characteristics of the effluent solids content, as observed from Figure 6 thru Figure 8, was the fact that the concentration of solids leaving the filter was not a direct function of the solids entering the filter. Generally the effluent should contain little solids if the filter system could be loaded within the range of allowable loading rate, despite the variation of influent solids concentration.

Because of the filter upflow nature and its long retention time, large quantities of solids could easily settle to the bottom of the filter where they have to be removed. However, no sludge was wasted during the entire 30 weeks operation. And no plugging was experience.

(E). Iron and Zinc Removal. Earlier study of DeWalle and Chian has indicated that the anaerobic filter has a high degree of metal reduction ability (14). Fe(III) and Zn(II) removals were tested for the present study and the results are shown in table 4.

According to table 4, all of the three phases achieved more than 92% reduction of Fe(III) and Zn(II). The highest concentrations of Fe(III) and Zn(II) in the leachate feed were 115.3 mg/l and 10.4 mg/l, respectively. With these metal concentrations, the AnSBF was able to remove 98.7% Fe(III) and 98.5% Zn(II) without influencing the filter ability on COD and BOD degradation. As a result, the final effluent of the AnSBF contained only 1.45 mg/l of Fe(III) and 0.15 mg/l of Zn(II).

Table 4 also indicates that filter 1 is more capable of reducing the metal ions than filter 2 and it always produced more than 87% metals removal. On the other hand, filter 2 can remove, at the most, 50% of metal ions remaining in the effluent of filter 1.

No metal toxicity was found because metal additions were increased gradually in order to allow the filter to acclimate to the new environment. However, the metal removal tests which were conducted by the present study were performed under the concentrations less than those reported by DeWalle and Chian. They stated that even with Fe(III) and Zn(II) concentrations as high as 430 mg/l and 16 mg/l, respectively, no significant effect on the anaerobic filter performance was observed.

(F). Gas Production. The anaerobic filter ability to generate a useable methane gas is one of the major reasons why it is being considered for full-scale use. Gas production as a function of organic loading is shown in Figure 9. It can be seen in Figure 9 that the gas production reached a steady-state condition in phase I, 62 days after the system start-up. Since phase I was operated under the highest organic loading, accordingly it produced the highest amount of gas at 22.3 liters per day. Phase III produced a slightly higher amount of gas than phase II (13.8 to 13.1 liters per day) even though phase II had a higher organic loading rate. This might be explained because phase III was conducted under a longer reactor retention time (16.7 days vs 9.5 days). The information used for the above discussion was provided in table 5.

The percentage of methane measured in the gas is given in table 5 and by multiplying the gas production by this percentage the amount of methane produced is obtained. And they are 16.99 liters per day, 10.72 liters per day, 10.80 liters per day for

*
Table 4. Metal Removal

Parameters	Phase I		Phase II		Phase III	
	Filter #1	Filter #2	Filter #1	Filter #2	Filter #1	Filter #2
A. Organic Loading,						
1b COD/10 ³ ft ³ day	141.8	93.9	72.2	4.90	60.7	5.80
1b BOD/10 ³ ft ³ day	106.7	60.0	53.4	3.90	47.2	1.40
B. Retention Time, days	8.0		17.5		8.7	
C. Influent Conc., §						
Fe ⁺³ , mg/l	32.4	2.37	30.1	3.71	115.3	1.95
Zn ⁺² , mg/l	0.85	0.09	0.71	0.04	10.4	0.19
D. Effluent Conc., §						
Fe ⁺³ , mg/l	2.37	2.16	3.71	2.37	1.95	1.45
% Fe ⁺³ Removal	92.7	28.0	87.7	36.1	98.3	26.3
% Overall Removal	93.3		92.1		98.7	
Zn ⁺² , mg/l	0.09	0.06	0.04	0.02	0.19	0.15
% Zn ⁺² Removal	89.2	32.9	94.3	50.0	98.1	26.9
% Overall Removal	92.8		97.1		98.5	

* Average of last 30 day data
§ Soluble Fe⁺³ and Zn⁺²

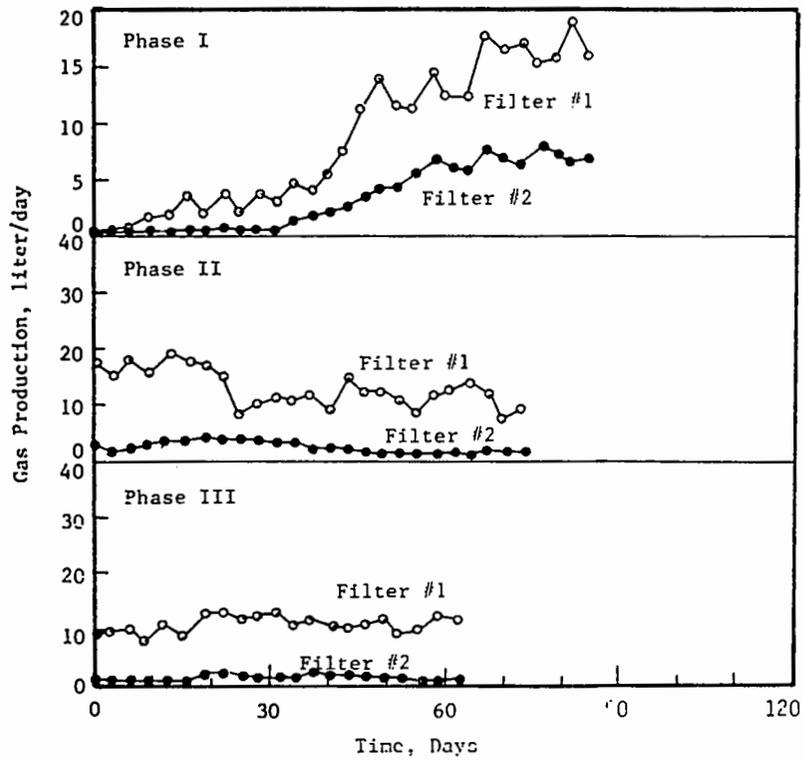


Figure 9 . Gas Production for Filters 1 and 2 Under Phases 1, 2, and 3 Operating Conditions

phases I, II, and III, respectively.

The theoretical methane production was also calculated and reported in table 5. It is apparent from table 5 that the theoretical methane production is slightly higher than the total measured methane production in all three phases. This difference was attributed to the utilization of organics for biological solids production and the removal of organics by the processes other than biological means.

Finally, in table 5 the amount of methane produced in cubic feet per pound of COD removed was calculated. The results show that the methane production was within the range of 5.01- 5.84 ft³ per lb COD removed, depending upon the organic loading and/or the reactor retention time employed.

CONCLUSIONS

The advantages inherent in anaerobic submerged filter system coupled with the amenability of this process in its application to leachate stabilization suggests that it is worthy of consideration as a basis for full scale leachate treatment facilities. The AnSBF is well suited to handling the large organic loads that often characterize leachates, particularly leachates that are discharged from young landfills. More than 96% COD and BOD can be removed from a high-strength acidic leachate wastewater, according to the present study, if the organic loading is controlled not in excess of 75.2 lb COD/1000 ft³-day or 56.4 lb BOD/1000 ft³-day with a reactor retention time of 16.7 days. The hydraulic retention time could be reduced to 9.5 days when the organic loading was decreased down to 62.3 lb COD/1000 ft³-day or 48.5 lb BOD/1000 ft³-day.

Metal removal efficiency is high when the influent concentrations of Fe(III) and Zn(II) are less than 115.3 mg/l and 10.4 mg/l, respectively. In addition to the efficient treatment of soluble COD, BOD, and heavy metals, the anaerobic filter requires no effluent recycle or sludge return. Also, the filter is able to reduce suspended and volatile solids in the leachate by over 90% because of its long retention time. Since the active biomass remains in the filter at all times, the system is ideally suited for the intermittent waste that a sanitary landfill might present.

The AnSBF also produces useable methane gas which could be used to heat the reactors. The volume of methane production is between 5.01 and 5.84 ft³/lb COD removed, depending upon the organic loading condition and the reactor retention time. In order to maintain the efficiencies mentioned above, the filter system has been heated at 36.5-38.3°C.

Table 5. Gas Production *

Parameters	Phase I		Phase II		Phase III	
	Filter #1 (A)	Filter #2 (B)	Filter #1 (A)	Filter #2 (B)	Filter #1 (A)	Filter #2 (B)
1. Temperature, °C	37.1	38.3	36.9	37.4	36.5	36.8
2. Organic Loading lb COD/10 ³ ft ³ day	141.8	93.9	72.2	4.90	60.70	5.80
lb BOD/10 ³ ft ³ day	106.7	60.0	53.4	3.90	47.20	1.40
3. Gas Production per filter, l/day	15.8	6.50	11.80	1.30	12.50	1.35
4. Total Gas Production, l/day, (3-A + 3-B)	22.3		13.1		13.8	
5. Measured % CH ₄	76.2	76.6	83.0	72.2	78.8	71.1
6. Measured Gas Production, l/day (3 x 5)	12.0	4.99	9.79	0.93	9.85	0.95
7. Total Measured CH ₄ Production, l/day (6-A + 6-B)	16.99		10.72		10.80	
8. Measured % CH ₄ , (7/4)	76.32		81.8		78.32	
9. Theoretical CH ₄ Production, l/day	19.45		12.5		11.40	
10. Theoretical % CH ₄ , (9/4)	87.2		86.8		85.79	
11. Measured: Theoretical (8/10)	0.874		0.942		0.912	
12. CH ₄ Produced/ lb of COD Removed, cu ft	5.84		5.01		5.78	

* Average of last 30 day data

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ENERGY RECOVERY FROM PRETREATMENT OF INDUSTRIAL
WASTES IN THE ANAEROBIC FLUIDIZED BED PROCESS

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INTRODUCTION

Due to the increasing cost of energy, there is an accelerated interest in using the anaerobic biological process for treatment of medium and high strength industrial wastewaters, therefore accomplishing both energy recovery and pollution control.

The fixed-film fluidized bed biological reactor utilizes the fluidized media to provide a large surface area for bacterial attachment and growth. A high biomass concentration can thus be retained in the reactor without biomass recycling. Dorr-Oliver's Anitron System™ employs the fluidized bed process for the anaerobic treatment of industrial wastewaters (Figure 1). The efficiency of the Anitron System has been demonstrated in many pilot scale operations, and a full scale plant is being constructed for treatment of soy protein processing wastewater (1). Further details of the system have been presented elsewhere (1,2). Two pilot plant studies involving anaerobic fluidized bed treatment of a dairy wastewater will be presented here.

The permeate from ultrafiltration of cheese whey for

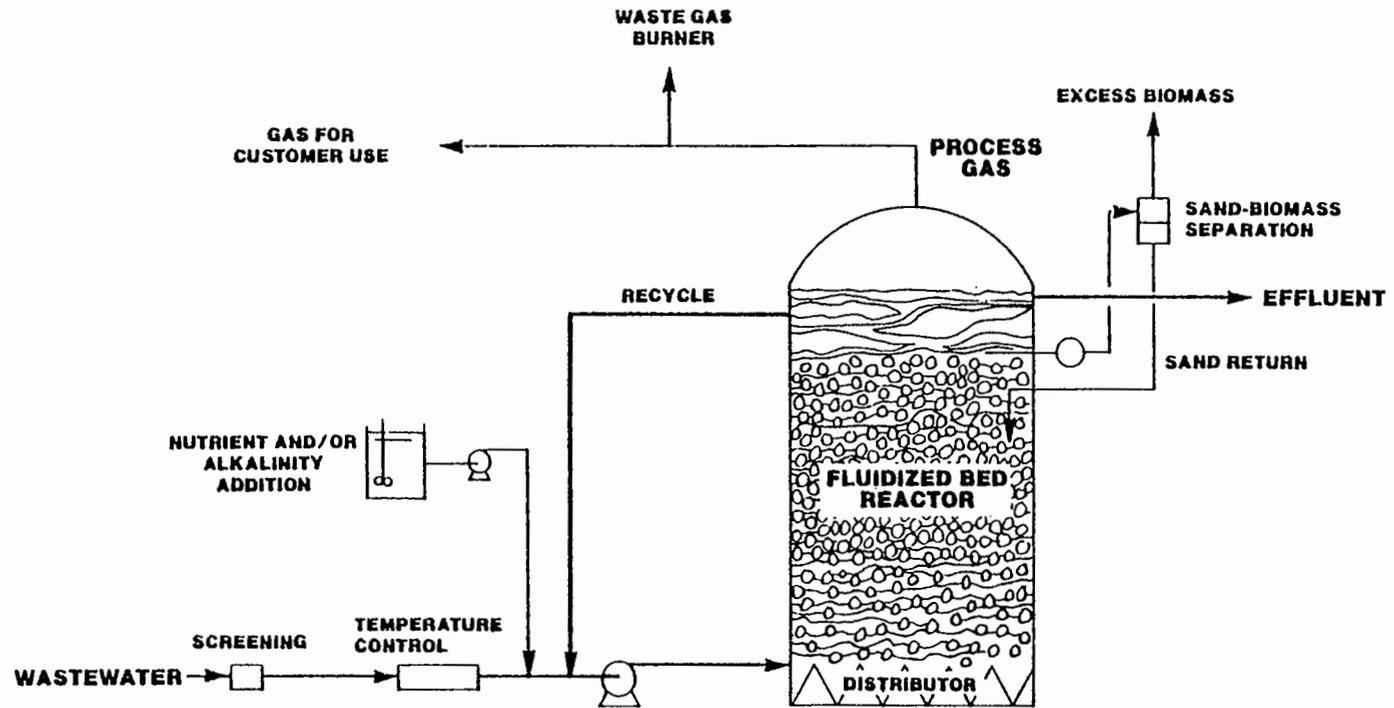


Figure 1. Anitron System Process Schematic

protein recovery has a 5-day biochemical oxygen demand (BOD₅) of approximately 40,000 mg/l, making it a severe pollutant if discharged without treatment (3). However, the high organic content of whey permeate also makes it an attractive substrate for energy recovery through anaerobic treatment.

The major component of whey permeate is lactose. Anaerobic degradation of lactose involves first the hydrolysis to glucose and galactose from which the volatile acids are produced by the acid formers. Conversion of the volatile acids to methane is thought to be through the formation of acetic acid, although other pathways, such as reduction of carbon dioxide and hydrogen, are also likely to occur (4). The methane bacteria have lower yields and growth rates and are more sensitive to environmental conditions than the acid-forming bacteria, thus methane fermentation is usually considered the rate-limiting step in the anaerobic treatment of soluble substrates, such as lactose. By providing a fluidized bed reactor, so that high concentrations of methane bacteria and acid bacteria can be maintained in the reactor, the rate of substrate removal and energy production can be enhanced, resulting in a high rate system.

Since the two groups of bacteria (i.e., the acid forming and methane forming species) differ widely with respect to their physiology and nutrient requirements, a separation of the acid phase from the methane phase by employing two reactors, each provided with optimal environmental conditions may enhance the rate of substrate removal allowing a reduction in the total reactor volume (5,6,7). This two-stage configuration may also increase the stability of the anaerobic process. The Anitron System is particularly suited to stage-wise operation because no unit process for suspended solids separation is required between stages.

This paper will report on the treatment of whey permeate in two pilot scale fluidized bed reactors operating in series and parallel configurations. The objectives of the study were:

- (1) to demonstrate the capability of the Anitron System to achieve high organic removal and methane production rates,
- (2) to evaluate the effect of feed concentration on the system performance, and
- (3) to determine the technical and economic advantages of a two-phase anaerobic fluidized bed system in comparison to a single-phase system.

OPTIMUM PROCESS DESIGN CONCEPT FOR THE ANITRON SYSTEM

Volumetric loading rate is commonly used in the design of attached growth systems such as trickling filters and rotating biological contactors, in which it is difficult to determine the reactor biomass concentration. The Anitron System can be designed using the same volumetric loading approach.

The volumetric loading (VL) rate to a system is defined as:

$$VL = \frac{Q S_o}{V} = \frac{S_o}{t} \quad (1)$$

and the volumetric removal (VR) rate which is a measure of the process performance is defined as:

$$VR = \frac{Q(S_o - S)}{V} = \frac{S_o - S}{t} \quad (2)$$

where,

Q = feed flow rate, $\frac{\text{volume}}{\text{time}}$,

S_o = feed substrate concentration, $\frac{\text{mass}}{\text{volume}}$,

S = effluent substrate concentration, $\frac{\text{mass}}{\text{volume}}$,

V = fluidized bed volume, volume,

t = reactor hydraulic retention time, time.

Optimal design of a fluidized bed reactor involves the evaluation of the process performance and the determination of the volumetric removal rates at various loading rates. The volumetric loading rate which meets the effluent quality requirements and results in a high removal rate is the choice for optimal design (Figure 2). Other factors, such as the process stability and the methane production rate should also be taken into consideration in determining the actual design loading.

PILOT PLANT FACILITIES AND OPERATION

The two pilot plants used were skid mounted, self-contained units, each having a 16.2 cm (6.4 in) diameter, 3.06 m (10 ft) high PVC fluidized bed reactor. Additional pilot plant components included:

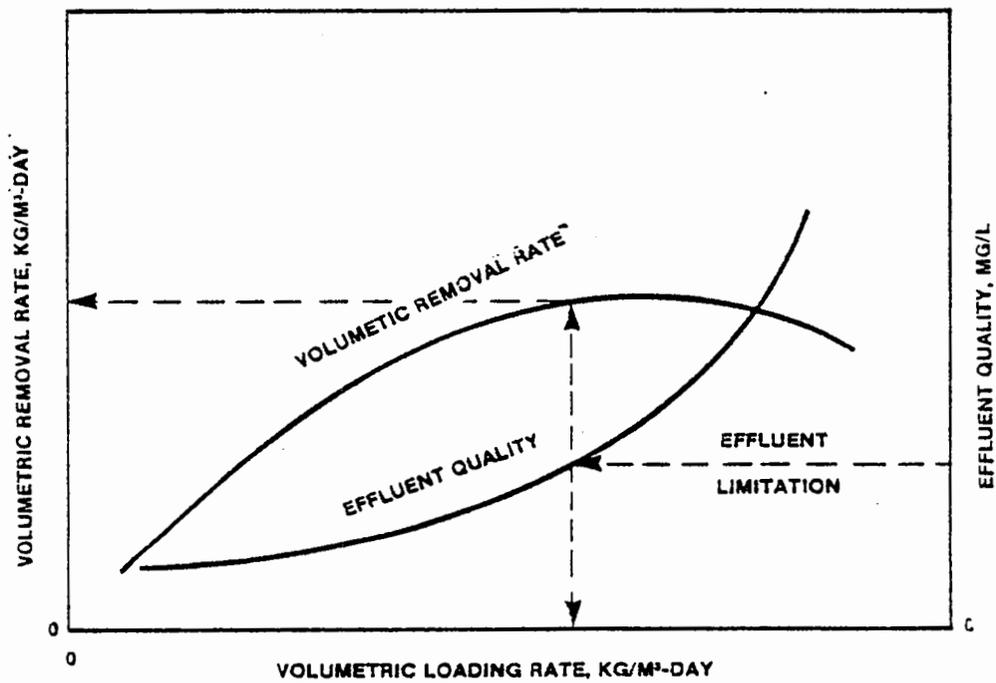


Figure 2. Approach to Preliminary Design of Anitron System Fluidized Bed Reactors

- a refrigerated feed tank, feed and recycle pumps,
- an electrical heat exchanger for automatic temperature control,
- a circulation pump and spray nozzle for foam control,
- a wet test meter for gas flow measurement,
- an in-line pH probe for continuous pH monitoring,
- a gas-liquid separator for removing the entrained gas from the effluent, and
- sampling ports for sludge wasting and bed solids sampling.

The pilot plant schematic is shown in Figure 3.

Start-up and Parallel Operation

The two pilot plant reactors (AN-1 and AN-2) were started up in parallel using supernatant from a municipal sludge digester as a source of methane seed organisms. Both units contained sand with a median particle size of 0.5 mm. The feed to the Anitron System reactors was prepared on-site by ultrafiltration of a 24 percent solution of sweet whey powder. The permeate was then diluted to the design concentration with dechlorinated tap water; approximately 10,000 mg/l of chemical oxygen demand (COD) for AN-1 and 30,000 mg/l for AN-2. Upon start-up of each reactor, temperature, pH, bed expansion, and recycle flow rate were monitored daily. The reactors were controlled under the same operating conditions (Table I) except for the feed concentration, the volumetric loading rate, and the degree of bed expansion allowed due to biomass growth. The volumetric loading rate was controlled at various levels by adjusting the feed rate to the reactors. Details of the pilot plant operation and the analytical procedures have been discussed and presented elsewhere (1).

Series Operation

In order to determine the effect of phase separation on system performance the two pilot plants were coupled together with reactor AN-2 being the first phase (acid formation) of the anaerobic process and reactor AN-1, the second phase (methane formation). The feed COD concentration to the coupled system was 10,000 mg/l. A refrigerated holding tank was installed between the two reactors to help balance the flow rate from AN-2 to AN-1. Operating conditions for both units (Table II) were controlled the same as under parallel operation except the pH in reactor AN-2 was maintained

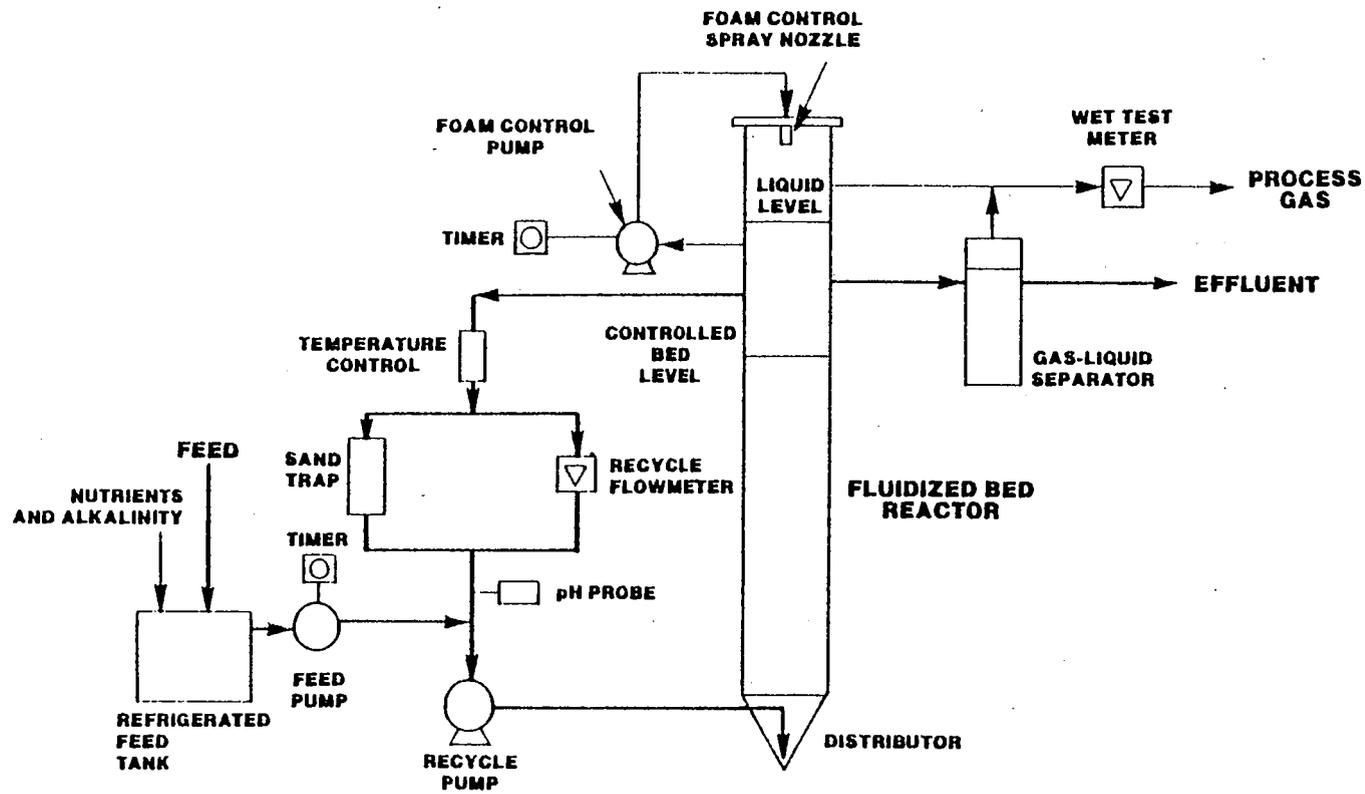


Figure 3. Anitron System Pilot Plant Schematic

TABLE I. Anitron System Operating Conditions
During Treatment of Whey Permeate:
Parallel Operation

Operating Conditions	AN-1	AN-2
Mean Sand Size, mm	0.5	0.5
Approximate Feed COD, mg/l	10,000	30,000
Hydraulic Loading Rate, m/min	0.41-0.55	0.41-0.55
Controlled Bed Expansion, %	90-110	40-60
pH Range	6.7-7.2	6.7-7.2
Temperature, °C	30-35	30-35

Note: 1 mm = 0.0394 in.
1 m/min = 24.5 gpm/ft²

TABLE II. Anitron System Operating Conditions
 During Treatment of Whey Permeate:
 Series Operation

Operating Conditions	Acid Forming Phase (AN-2)	Methane Forming Phase (AN-1)
Mean Sand Size, mm	0.5	0.5
Approximate Feed COD, mg/l	10,000	AN-2 Effluent
Hydraulic Loading Rate, m/min	0.41-0.55	0.41-0.55
Controlled Bed Expansion, %	40-60	90-110
pH Range	5.7-6.2	6.7-7.2
Temperature, °C	30-35	30-35

Note: 1 mm = 0.0394 in.
 1 m/min = 24.5 gpm/ft²

between 5.7 to 6.2 instead of 6.7 to 7.2. Lowering the pH is believed to inhibit the growth of the methane bacteria, thus promoting the growth of acid formers (5,6).

RESULTS AND DISCUSSION

The experimental results presented here include:

- (1) evaluation of the single phase Anitron System operating at various loading rates,
- (2) assessment of the effect of feed concentration on process performance, and
- (3) evaluation of the results from two-phase operation of the Anitron System and comparison to the single-phase results.

Process Performance of the Single-Phase Anitron System

The volumetric loading rate to AN-1 was increased gradually, and the performance was evaluated at the desired volumetric loading rate. Feed concentration to the reactor was maintained as close to 10,000 mg/l as possible. The operating conditions were presented in Table I.

Figure 4 shows the effluent COD and the corresponding COD removal as a function of volumetric loading rate. Only a slight increase in effluent COD and a slight decrease in COD removal were detected when the volumetric loading rate was increased from 8 to 24 kg COD/m³-day (0.5 to 1.5 lb COD/ft³-day).

BOD₅ values were calculated from the correlation between BOD₅ and COD, and excellent correlations were obtained for both influent and effluent values. Removals for BOD₅ were generally higher than COD removals.

Effluent suspended solids (SS) concentration remained pretty much the same at various loading rates (Figure 5). An average of 361 mg/l SS was obtained for all the experimental runs. However, an increase in percent SS removal was observed as organic loadings increased. This could be attributed to a higher feed SS concentration during high volumetric loading rates. The results indicate that influent SS concentration up to 2,500 mg/l can become hydrolyzed in the Anitron System reactor and the effluent SS value does not appear to be affected by the feed SS concentration.

The volumetric removal rate, as defined previously in Equation (2), is plotted against the volumetric loading rate

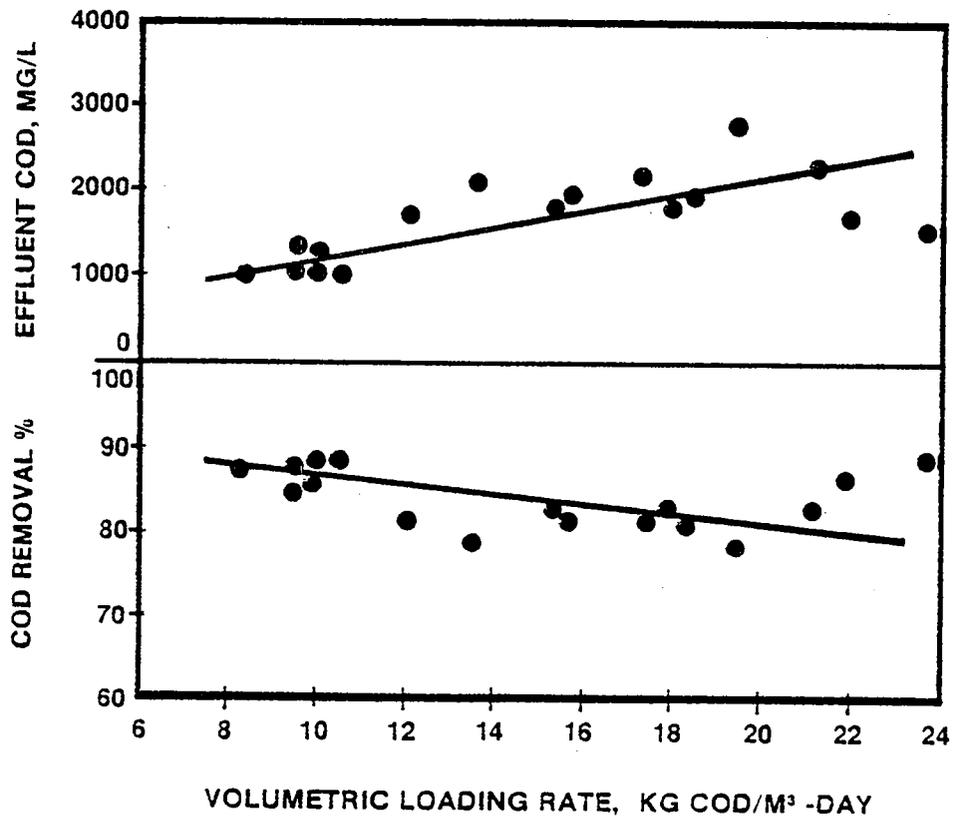


Figure 4. Performance of Single-Phase Anitron Reactor at Various Volumetric Loading Rates.

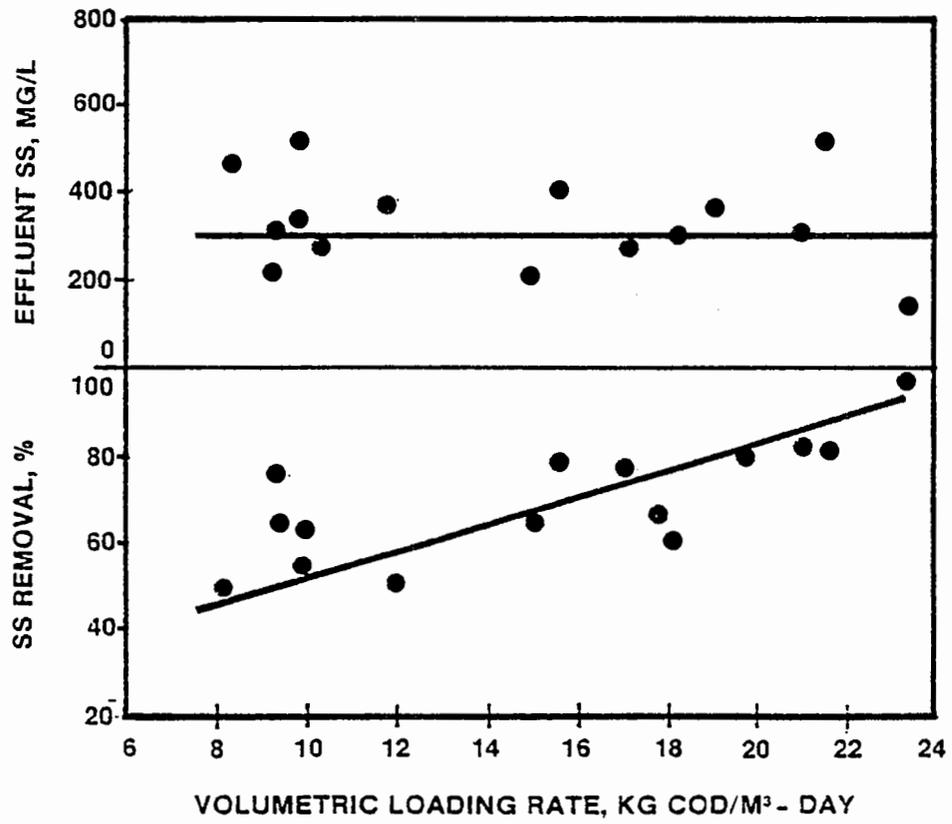


Figure 5. Effluent SS and SS Removal for Single-Phase Anitron Reactor at Various Loading Rates.

for the experimental data obtained and is shown in Figure 6. The relation appears to be linear, and can be expressed as:

$$VR = 0.82 (VL) \quad (3)$$

$$\text{For } 8 < VL < 24 \text{ kg COD/m}^3\text{-day}$$

By comparing the results in Figure 6 to the theoretical design curve (Figure 2), it is evident that the maximum volumetric removal rate for the Anitron System had not yet been reached even at a VL value of 24 kg COD/m³-day. Increasing the volumetric loading rate will further reduce the reactor volume required and increase the energy production rate per unit volume. The energy production rate in kcal/m³-day can be calculated based on the experimental methane yield of 0.3 m³ CH₄ produced per kg of COD removed and the heating value of methane at 8540 kcal/m³. Its relationship to volumetric loading is also shown in Figure 6.

The experimental data discussed above are summarized in Table III based on the system or reactor hydraulic retention time (HRT). Mean influent, effluent, and removal values for COD, BOD₅, and SS are tabulated together with effluent volatile acids (VA) and volumetric removal rates. A slight decrease in COD and BOD removal and a gradual increase in SS removal as volumetric loading increases, are consistent with previous discussions.

The average observed biogas production was 0.510 m³/kg COD removed (at STP) at a volumetric loading of 16.2 kg COD/m³-day, and 0.464 m³/kg COD at 19.3 kg COD/m³-day. With an average methane content of 60 percent in the biogas, the corresponding methane yields are respectively 0.306 and 0.278 m³/kg COD. These results compare with values from 0.204 to 0.335 m³/kg COD for treatment of whey (8), and 0.274 m³/kg COD for treatment of whey permeate (1). The theoretical methane production is 0.35 m³/kg COD at STP (5.6 ft³/lb).

The observed biomass yield determined for AN-1 using the COD mass balancing approach (2) was found to be 0.15 g VSS/g COD. This value is lower than the 0.24 g VSS/g COD value determined in the previous study at a volumetric loading of 6.5 kg COD/m³-day (1). This discrepancy could be just an inherent error from the COD balancing method, not because of the loading difference.

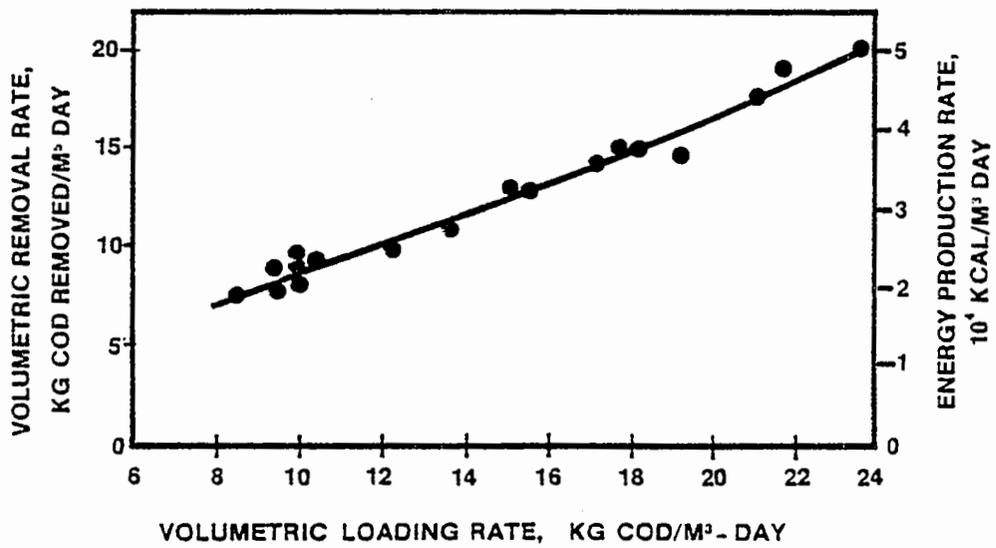


Figure 6. Volumetric Removal Rates and Energy Production Rates as a Function of Volumetric Loadings.

Table III. Anitron System Performance Results During Treatment of Whey Permeate: Single Phase Operation

Parameters	AN-1		
System Hydraulic Retention Time, h	41.1	29.4	25.7
Reactor Hydraulic Retention Time, h ^{a)}	23.3	16.6	14.5
Volumetric Loading, kg COD/m ³ -day ^{a)}			
Mean	9.6	16.2	19.3
Range	8.3-10.4	15.2-17.3	18.0-23.8
Mean Influent Value, mg/l			
COD	9374	13159	12297
BOD ₅ ^{b)}	5457	7370	6933
SS	844	1193	2481
Mean Effluent Value, mg/l			
COD	1141	1954	2015
BOD ₅ ^{b)}	560	836	857
SS	358	416	309
VA (as acetic)	294	565	585
Mean Removal, %			
COD	87.8	85.2	83.6
BOD ₅ ^{b)}	89.7	88.7	87.6
SS	57.6	65.1	87.5
Mean Volumetric Removal Rate, ^{a)} kg COD removed/m ³ -day	8.4	13.8	16.1
Biogas Yield at STP m ³ produced/kg COD removed	-	0.510	0.464
Methane Yield at STP m ³ produced/kg COD removed	-	0.306	0.278

a) Based on fluidized bed volume.

b) BOD₅ values based on correlation developed between influent and effluent COD and BOD₅.

c) 1 l = 0.264 gal
1 kg/m³-day = 62.5 lb/1000 ft³-day
1 m³/kg = 16 ft³/lb

Effect of Feed Concentration on Anitron System Process Performance

The effect of feed strength on system performance was evaluated by comparing the results from the parallel operation of the two Anitron units, each receiving different feed concentrations. The feed to AN-1 had a COD of approximately 10,000 mg/l and the feed to AN-2, 30,000 mg/l (Table I). The performance results for the two units operating at the same mean volumetric loading rates are shown in Table IV.

The process performance of AN-2 appeared inferior to that of AN-1 at the same volumetric loading of about 19.5 kg COD/m³-day: a 76.4% COD and 77.5% BOD removal for AN-2 as compared to 83.6% COD and 87.6% BOD for AN-1 (Table IV). This lower effluent quality can be attributed to a combination of the higher volatile acids and the higher SS concentration in the effluent from the reactor.

The results derived during performance assessment of the single-phase Anitron System indicate that the influent SS do not contribute to the effluent SS concentration. It is more likely that the suspended solids in the AN-2 effluent are due to biomass shearing or sloughing from the sand media and/or biomass growth in the liquid phase. If shearing or sloughing was the major contributor to the problem, we would expect that the effluent suspended solids from reactor AN-1 would be greater than that from AN-2. However, this was not the case. Reactor AN-1 was controlled at greater percent fluidized bed expansion than AN-2 (Table I), and therefore the biofilm thickness will be greater in this reactor (9). Shearing or sloughing should be enhanced by thicker biofilms. The possibility that biomass growth in the liquid phase accounts for the high effluent SS concentration is supported by the fact that AN-2 had a system hydraulic retention time (HRT) of 77.1 hours, which is three times longer than the system HRT for AN-1 (25.7 h). The system HRT becomes the system solid retention time (SRT) for the biomass in the liquid phase in a complete-mix reactor. The Anitron System is considered complete-mix when operating at high recycle ratios. SRT is a measure of the bacterial growth rate and microbes with growth rates greater than the reciprocal of the system SRT should grow and proliferate. Under this premise, anaerobic organisms with growth rates greater than 0.31 days⁻¹ (reciprocal of 77.1 h) will exist in the liquid phase of reactor AN-2, thereby contributing to the effluent suspended solids concentration.

Table IV. Anitron System Performance Results During Treatment of Whey Permeate: Effect of Feed Concentration

Parameters	AN-1	AN-2
System Hydraulic Retention Time, h	25.7	77.1
Reactor Hydraulic Retention Time, h ^{a)}	14.5	43.6
Volumetric Loading, kg COD/m ³ -day ^{a)}		
Mean	19.3	19.7
Range	18.0-23.8	15.9-21.5
Mean Influent Value, mg/l		
COD	12297	35408
BOD ₅ ^{b)}	6933	18650
SS	2481	2601
Mean Effluent Value, mg/l		
COD	2015	8356
BOD ₅ ^{b)}	857	4195
SS	309	2295
VA (as acetic)	505	2014
Mean Removal, %		
COD	83.6	76.4
BOD ₅ ^{b)}	87.6	77.5
SS	87.5	14.5
Mean Volumetric Removal Rate, ^{a)} kg COD/m ³ -day	16.1	15.1
Biogas Yield at STP m ³ produced/kg COD removed	0.464	0.452
Methane Yield at STP m ³ produced/kg COD removed	0.278	0.272

a) Based on fluidized bed volume.

b) BOD₅ values based on correlation developed between influent and effluent COD and BOD₅.

c) 1 l = 0.264 gal
1 kg/m³-day = 62.5 lb/1000 ft³-day
1 m³/kg = 16 ft³/lb

If the above discussion on the effect of the feed concentration is true, then in the treatment of whey permeate there exists a maximum system HRT under which the anaerobic fluidized bed process can operate without having a suspended solids accumulation problem in the reactor. This critical system HRT was found to be between 41 and 77 hours, which corresponds to a minimum growth rate of 0.31 to 0.59 days⁻¹. The lower limit of 41 hours for the critical system HRT was obtained from the experimental data derived during assessment of the single-phase Anitron System (Table III). For a design loading rate of 16 kg COD/m³-day, the corresponding limiting feed concentration is in between 15,000 mg/l and 30,000 mg/l of COD.

The negative impact of high substrate concentration on Anitron System performance was further verified in subsequent pilot plant operation at a feed concentration of 60,000 mg/l.

The biogas and methane yields determined for units AN-1 and AN-2 were essentially the same (Table IV). This implies that feed concentration or hydraulic retention time has no effect on the value of these parameters.

Process Performance of the Two-Phase Anitron System

Following assessment of the single-phase Anitron System, the reactors were coupled together and the performance of the two-phase system was determined. In this study AN-2 was used as the acid phase reactor, and AN-1 the methane phase. Data were then derived under the operating conditions previously presented (Table II).

The experimental results for the two-phase system are shown in Table V. Although the mean volumetric loading to the first phase system (AN-2) was 21.1 kg COD/m³-day, the total system volumetric loading was only 10.6 kg COD/m³-day. The acid phase reactor removed an average 53.2% of COD and the methane phase reactor removed an additional 40.7% to give a total of 93.9% of COD removal for the entire system. The BOD removal for the two-stage system was also high, over 97%. Both COD and BOD removals in the two-phase system are higher than in the single stage Anitron System (Table III) operating at a similar loading level (9.6 kg COD/m³-day) and system SRT (41.1 h).

Preliminary results on the gas composition have shown a 30% methane content for the gas generated in the acid phase reactor and 87% for the methane phase. Work is continuing to determine the methane production for the two-stage system.

Table V. Anitron System Performance Results During Treatment of Whey Permeate:
Two-Phase Operation

Parameters	Acid Phase AN-2	Methane Phase AN-1	Total System
System Hydraulic Retention Time, h	25.3 ^{b)}	18.1	43.4
Reactor Hydraulic Retention Time, h ^{a)}	10.9	10.1	20.9
Volumetric Loading, kg COD/m ³ -day ^{a)}			
Mean	21.1	10.6	10.6
Range	17.6-26.7	3.3-14.6	8.3-13.4
Mean Influent Value, mg/l			
COD	9490	4441	9490
BOD	4620	2603	4620
TSS	352	728	352
Mean Effluent Value, mg/l			
COD	4441	579	579
BOD	2603	140	140
TSS	728	213	213
VA (as acetic)	1306	29	29
Mean Removal, %			
COD	53.2	87.0	93.9
BOD	43.7	94.6	97.1
TSS	-	-	38.1
Mean Volumetric Removal Rate, ^{a)} kg COD/m ³ -day			
	11.2	9.2	10.0

a) Based on fluidized bed volume.

b) Includes the holding tank volume.

c) 1.0 = 0.264 gal
1 kg/m³-day = 62.5 lb/1000 ft³-day
1 m³/kg = 16.05 ft³/lb

CONCLUSIONS

The conclusions derived from the results presented in this paper include the following:

1. High strength industrial wastewaters, such as whey permeate, can be effectively treated by the anaerobic fluidized bed process and the energy generated, recovered. COD and BOD₅ removals of over 80% can be obtained at volumetric loading rates over 24 kg COD/m³-day (1.5 lb COD/ft³-day). The observed methane production will be approximately 0.3 m³/kg COD at STP.
2. Parallel operation of two Anitron System reactors at two different levels of feed concentration revealed that there appears to exist a maximum system hydraulic retention time beyond which microbial solids begin to proliferate in the liquid phase of the system. In the treatment of whey permeate at a volumetric loading rate of 16 kg COD/m³-day, the maximum system HRT corresponds to a maximum feed COD concentration of approximately 15,000 mg/l.
3. Results to date indicate the superior performance of the two-phase operation over the single phase system at a volumetric loading of 10 kg COD/m³-day. Studies are under way to evaluate the two-phase performance at higher loading rates.

ACKNOWLEDGEMENTS

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PART XIV: PROCESS EVALUATION AND DESIGN

THE HYDRODYNAMIC EVALUATION OF A FIXED MEDIA BIOLOGICAL PROCESS

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INTRODUCTION

Two basic classifications of flow models for biological processes are presently accepted, plug or piston flow and complete mixing. The trickling filter can be modelled as a plug flow system and the activated sludge process can be designed as a complete mixed system. If a trickling filter and a complete mixed activated sludge system are combined together in a tank, then an interesting question arises on how the mixing characteristics will be changed.

A hydrodynamic evaluation was made to characterize the flow patterns and the mean detention time changes inherent to such a system. Tracer analyses were used during the course of the development of a fixed media process which utilized plastic tower packing, but unlike a trickling filter system, the packing was completely submerged. A surface aerator supplied oxygen to the system and provided mixing in the system by pumping action through a draft tube.

EXPERIMENTS

Three different reactors were used during the course of development of the fixed media process. The characteristics of the media used for this study are illustrated in Table I

Table I Physical Characteristics of Media

Packing	Nominal Size (cm)	Specific Surface Area (m^2/m^3)	Void Space (%)
Flexiring	1.6	345	92
Flexiring	3.8	132	96
Flexiring	8.9	92.1	97

Pilot Unit

This unit was equipped with a draft tube and a surface aerator as shown in Figure 1. The main component of the unit was the biological reactor, which was 2.4 m in diameter with a 3 m side-water depth. A draft tube 46 cm in diameter was located in the center of the tank. The tank was equipped with a circular cone hopper bottom where solids could settle. The total liquid volume of the tank was about 15.6 cu m without the fixed media, and about 14.7 cu m with the media. The liquid volume of the media zone was 12.2 cu m. A total of 13,800 8.9 cm flexirings was placed in the tank.

Degrittied raw domestic wastewater was introduced above the liquid level inside of the tank. The influent was thoroughly and instantly mixed with the tank liquid being pumped by the aerator. This produced a uniform organic loading on the tank surface area. The aerator was driven by a 1/2 hp motor (later changed to 3 hp) and rated at about 100 rpm. The aerator had 4 blades, each of which was 10 cm in width. The tip to tip diameter of the aerator was 47 cm.

16.5 Liter Unit

This unit was constructed to simulate the shape of the pilot unit. It was utilized to determine the amount of solids which accumulated on different sizes of media, and to develop a possible solid handling method. A total of 1,900 flexirings was placed in the media zone for 1.6 cm media operation, and a total of 177 flexirings was provided for 3.8 cm media operation. Both the 1.6 and 3.8 cm media were randomly packed, while a total of 15 units of media was specifically situated in the reactor for 8.9 cm media operation.

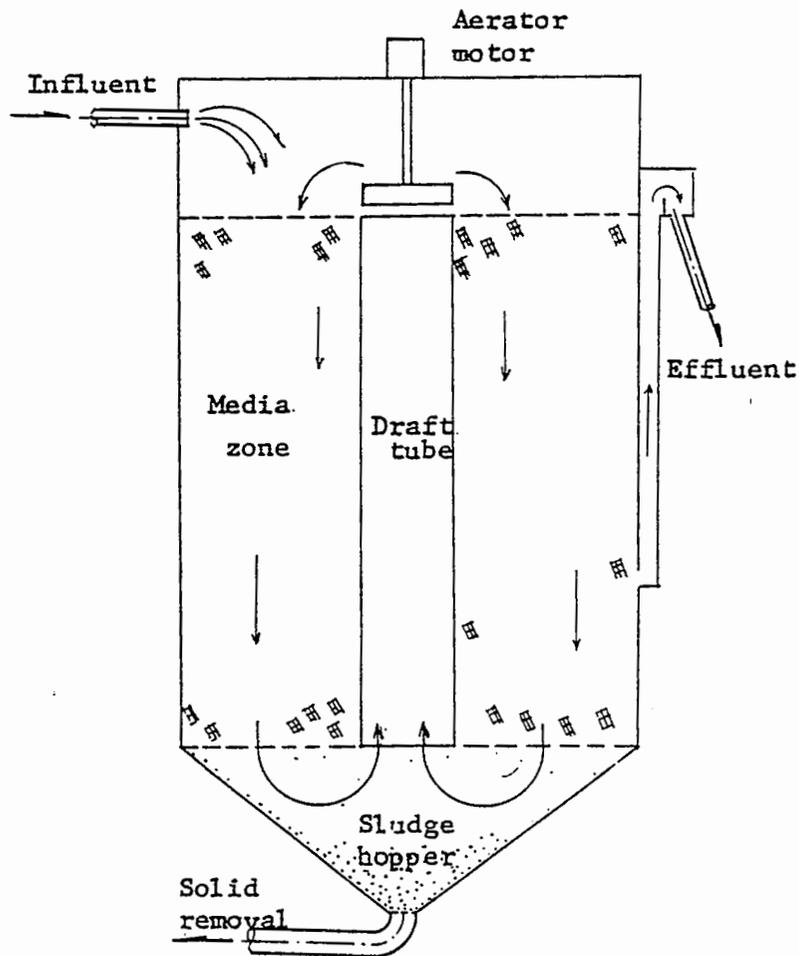


Figure 1. Pilot Unit Profile

The reactor volume was 16.5 l with the 1.6 and 3.8 cm media, however the volume was changed to 17.5 l with the 8.9 cm media. Two aerator assemblies manufactured by Virtis were provided. One assembly was located inside of the draft tube and the other exposed from the draft tube wall similar to the aerator in the pilot unit. Primary effluent was fed to the reactors.

4 Liter Units

In addition to the solids removal and media size problems, it was not known how many solids would be produced per unit time and how often these solids should be wasted. For these questions another investigation was made by using a series of 5-4 l reactors. Each reactor had the same configuration and was operated by feeding synthetic substrate made with skim milk and inorganic salts. The feed BOD concentration was 670 mg/l. All reactors had the same number of media, 550 1.6 cm flexirings. A bottom hopper was not provided in order to eliminate the problem of supporting the units.

Mixing Tests

It was expected that the flow patterns and the mean detention times of the reactors were changed with the growth of microbes on the surface of the media. Therefore, mixing studies were made by utilizing a flourometer or a specific chloride ion electrode.

Pulse dye or chloride input was made to observe exit age distribution. The time consumed to inject the dye or chloride into the reactors was always less than 1 second. Calibration of the measurement device with dye or chloride was made prior to the actual mixing tests. Since the flourometer reading scale ranged from zero to 100, the amount of dye was added to fall within the available scale. The expected dye reading (C_o) was computed based upon the assumption that the reactors were completely mixed. A temperature correction was made using the following formula prior to analyzing the data:

$$C_{20} = C_i e^{0.029(T-20)} \quad (1)$$

where T = temperature in °C

C_i = dye reading at temperature T, and

C_{20} = calibrated dye reading at 20°C

Sodium chloride was used for the chloride injection. The expected chloride concentration was also computed based upon the assumption that the reactors were completely mixed. To avoid the measurement error which would occur at low concentrations, more than 2,000 mg of NaCl was added into the reactor. Also, to avoid possible effects upon the microbial population, the chloride concentration was maintained less than 2,500 mg/l.

For the pilot and 16.5 l units without microbial growth, flourocein dye was used to characterize the mixing patterns in the reactors; while pontacyl pink B dye was utilized for the 16.5 l unit with microbial growth.

The mixing test results were analysed by the methods described by Levenspiel (1). All analytical procedures for water quality were done according to Standard Methods (2).

RESULTS

Pilot Unit

Figure 2 shows the exit dye readings for the pilot unit. Both with and without media systems approached complete mixing. The washout rate was estimated to be $-0.104/\text{hr}$, which is equivalent to a 9.6 hour mean detention time. About 20 percent of the total tank volume appeared to be dead space. By extending the washout curve to 4,000 minutes, 105 percent of the tracer dye was recovered. The maximum C_1/C_0 was 1.08. The dispersion number and Morrill Index (3) were 0.7837 and 10.99, respectively. The computed mean detention time utilizing the moments of distribution method (1) was 7.4 hours compared to the 9.6 hours previously mentioned.

16.5 Liter Unit

There were a number of dye tests conducted on this unit, but dye adsorption and mechanical problems limited the amount of usable data. Figure 3 illustrates the results obtained for the different flow rates and aerator speeds.

Generally speaking the mixing curves were skewed to the right which means that the system approached complete mixing. However, the nonlinearity of the curves indicates a series of complete mixing cells or a system approaching plug flow conditions. This possibility could not be confirmed since dye recovery ranged from 30 to 70 percent for these runs.

After about 4 months from the start-up of the reactor dye studies were made under actual operating conditions. Since fluorescein appeared to be adsorbed on the plexiglas (as mentioned previously), pontacyl pink B dye was used. The mixing curves (a) and (b) shown in Figure 4 were obtained with a heavy microbial growth, while curve (c) was

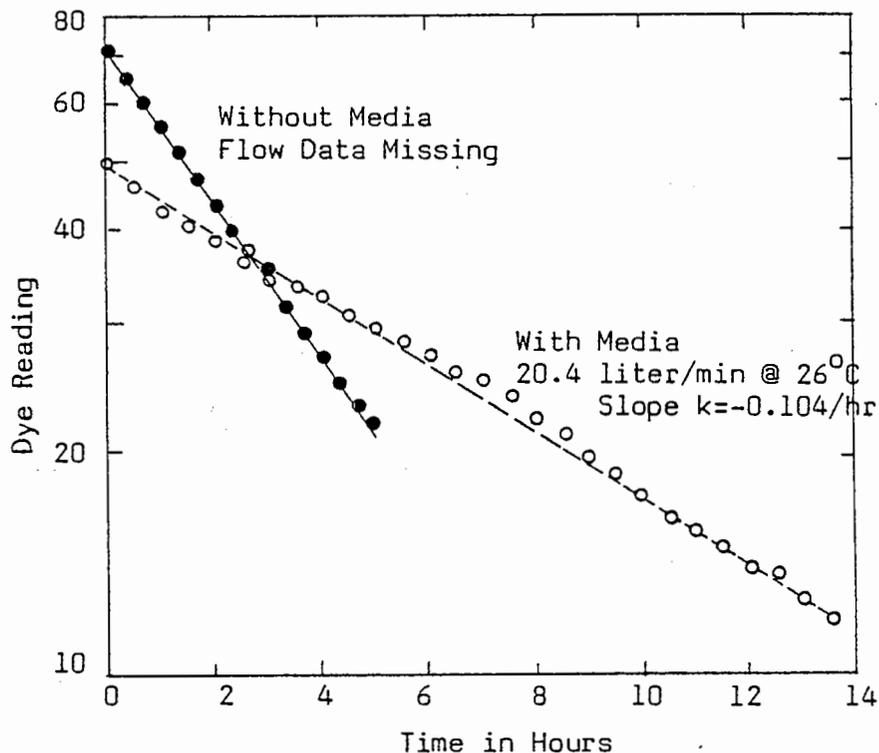


Figure 2. Washout Characteristics of Pilot Unit Without Microbial Growth

obtained with a smaller microbial population on the 3.8 cm media. The occurrence of the maximum dye concentration appeared at the middle of the curve with the higher microbial growth, while curve (c) would indicate a complete mixing pattern with some short circuiting.

Like fluorescein, pontacyl pink B dye recovery was bad. Later it was visually observed that some added dye stayed on the microbial floc surfaces.

4 Liter Unit

Table II is a summary of the mixing characteristics of the media system without microbial growth. Reactors 1 to 5 were put into operation simultaneously to investigate the solid accumulation rate. Reactor No. 1 was operated only for 12 hours, No. 2 for 24 hours, No. 3 for 48 hours, No. 4 for 72 hours, and No. 5 for 96 hours. Mixing tests under

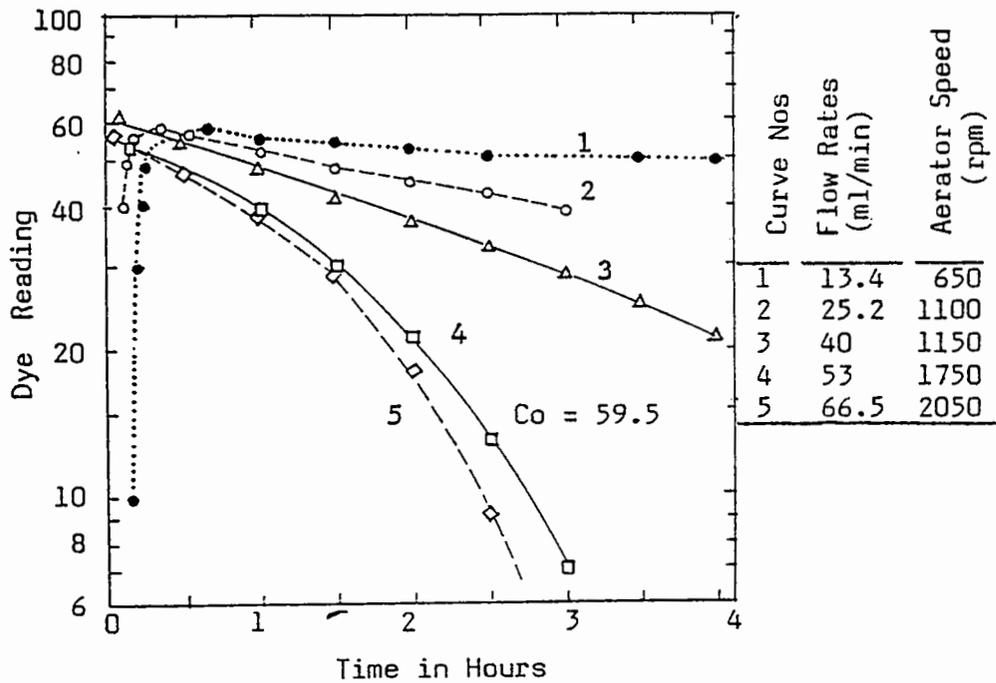
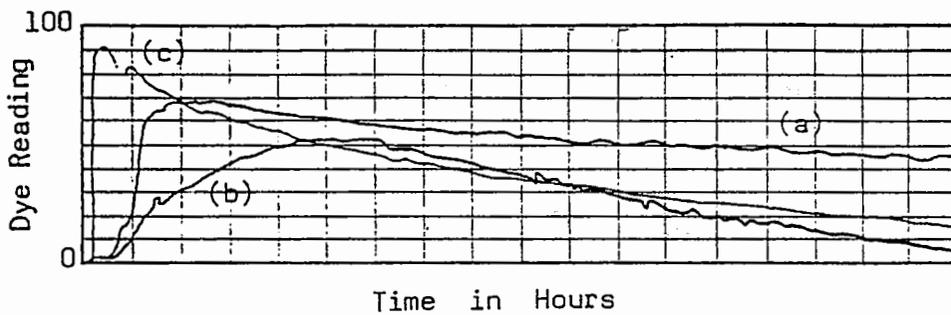


Figure 3. Washout Characteristics of 16.5 Liter Unit Without Microbial Growth



Curve Nos	Days from Start-up (days)	Flow Rate (ml/min)	Operating Temp (°C)	Media Size (cm)
(a)	116	15	25	1.6
(b)	120	88	25	1.6
(c)	130	51	22	3.8

Figure 4. Washout Characteristics of 16.5 Liter Unit With Microbial Growth

Table II Mixing Characteristics Without Microbial Growth - Nominal 4 Liter Unit

	Unit Numbers				
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
Flow Rates (1/hr)	0.91	0.93	0.99	0.99	0.95
Reactor Volumes (l)	3.94	3.98	4.0	3.93	3.92
	<u>By Regression Analyses</u>				
Theoretical Detention (hr)	4.33	4.28	4.04	3.97	4.13
Washout Rates 10k (hr ⁻¹)	2.34 (2.22-2.35)	2.42 (2.22-2.62)	2.53 (2.62-2.90)	2.74 (2.62-2.90)	2.60 (2.49-2.76)
Mean Detention Time (hr)	4.28 (4.26-4.52)	4.14 (3.81-4.52)	3.96 (3.45-4.52)	3.65 (3.45-3.81)	3.81 (3.62-4.02)
	<u>By Utilizing Moments of Distribution</u>				
Mean Detention Time (hr)	3.91	3.63	3.62	3.50	3.60
σ^2	1.0699	1.0877	1.0893	1.0845	1.0776
Dye Recovered (%)	96.1	91.5	90.2	89.3	90.2
Morrill Indices	23.0	32.6	45.3	-	43.0

microbial growth were conducted at 4 different times following start-up of the media operation with microbial growth: the 1st day test designates the mixing test conducted during hours 1 to 13, the 2nd day during hours 26 to 36.3, the 3rd day during hours 53 to 60, and the 4th day test during hours 73 to 84.5. Table III is a summary of the mixing test results, and Figure 5 shows the exit NaCl concentrations for reactor No. 5. Approximately 2,400 mg of NaCl was injected into the reactor without microbial growth operation and approximately 6,000 mg of NaCl was injected into the reactor during the actual operating conditions. It should be noted that ponding occurred in unit No. 5 on the 4th day of operation.

The washout rates for the 4 l reactors were obtained from regression analyses. The values in parentheses indicate the confidence limits at the 95 percent level. By computing areas of the tracer curves, about 89 to 96 percent of the total dye added was recovered.

The Morrill Indices ranged from 23 to 43. Since the variances were close to 1.0, it is apparent that the reactors approached ideal complete mixing flow patterns.

DISCUSSION

Mean Detention Time

Microbes and influent materials were accumulated on the media by adsorption, settling, flocculation, and growth. The solids were accumulated and compacted by the continuous growth. Eventually this will lead to anaerobic conditions unless the excessive solids are wasted. The extent to which these materials accumulated on the media is referred to as the maximum solids holding capacity as shown in Table IV. As the solids accumulated on the media, the detention time was changed as shown in Table III, and the mixing characteristics changed from complete mixing to plug flow as shown in Figure 4.

In this study, the exit age distribution was extensively used for analyzing mixing data, but the methods available are somewhat ambiguous. The mathematical formula given for an ideal complete mixing regime is:

Table III Mixing Characteristics With Microbial
Growth - Nominal 4 Liter Unit

	Unit Numbers				
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
Corrected Flow Rates (1/hr)	0.97	1.02	1.01	0.98	0.95
	<u>1st Day Operation</u>				
Mean Detention Time (hr)	3.43 (3.29-3.62)	3.62 (3.45-3.81)	3.67 (3.62-3.81)	3.62 (3.45-3.81)	3.81 (3.62-4.02)
	<u>2nd Day Operation</u>				
			3.15 (2.90-3.45)	3.62 (3.45-3.81)	4.16 (3.81-4.52)
	<u>3rd Day Operation</u>				
				2.48 (2.07-3.01)	2.82 (2.41-3.44)
	<u>4th Day Operation</u>				
					4.06 (3.80-4.52)

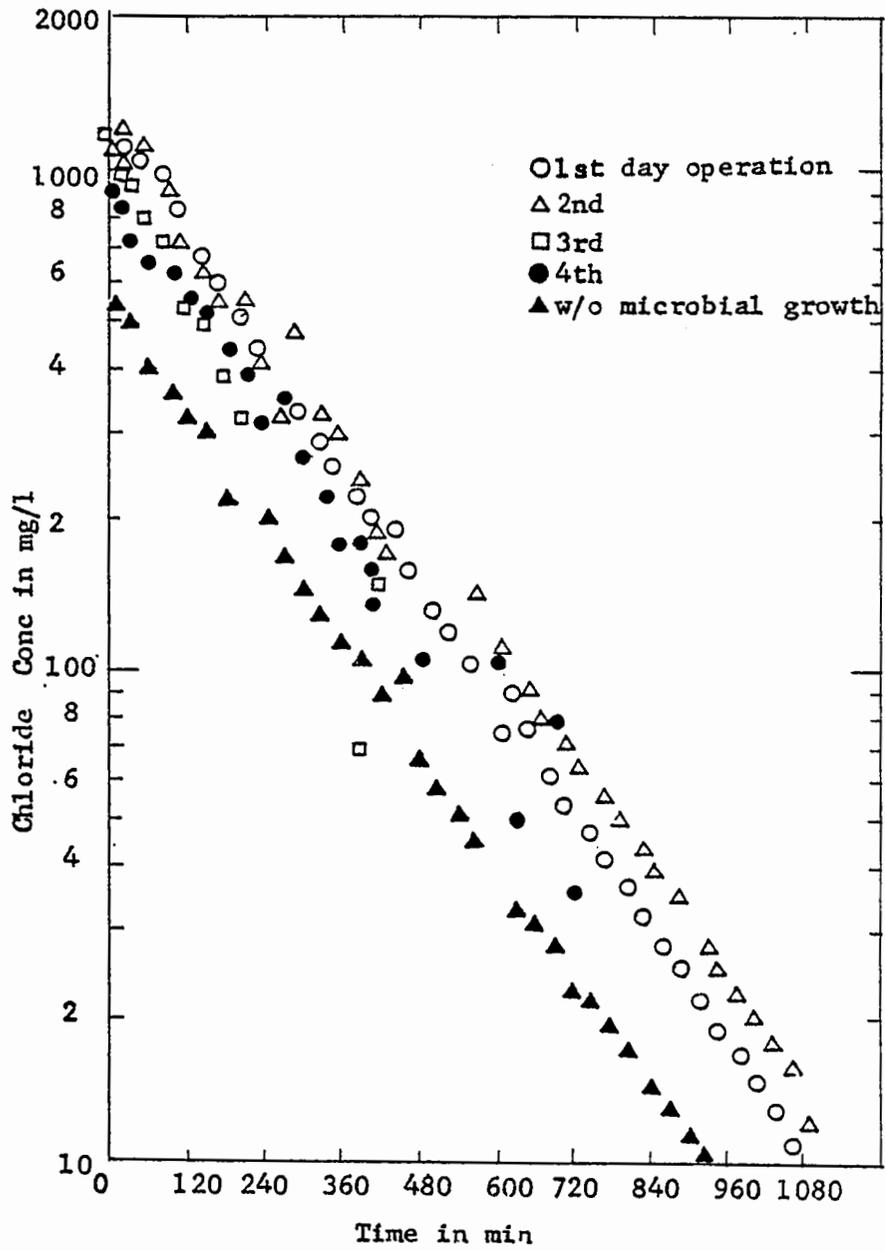


Figure 5. Washout Chloride Conc of Unit No. 5

$$C_i = C_o e^{-kt_i} \quad (2)$$

C_i = dye or chloride concentration at exit

C_o = initial dye or chloride reading when the reactor is completely mixed

k = washout rate per time = $1/\bar{t}$

\bar{t} = mean detention time

t_i = time

Table IV Estimated Maximum Solid Holding Capacity

Medium Size (cm)	Max MLSS (mg/l)	Solids Holding per Unit Area (g/sq m)	Solids on Media (%)
1.6	8,100	30.1	2 - 4
3.8	7,500	65.7	4 - 6
8.9	5,900	67.8	6 -17

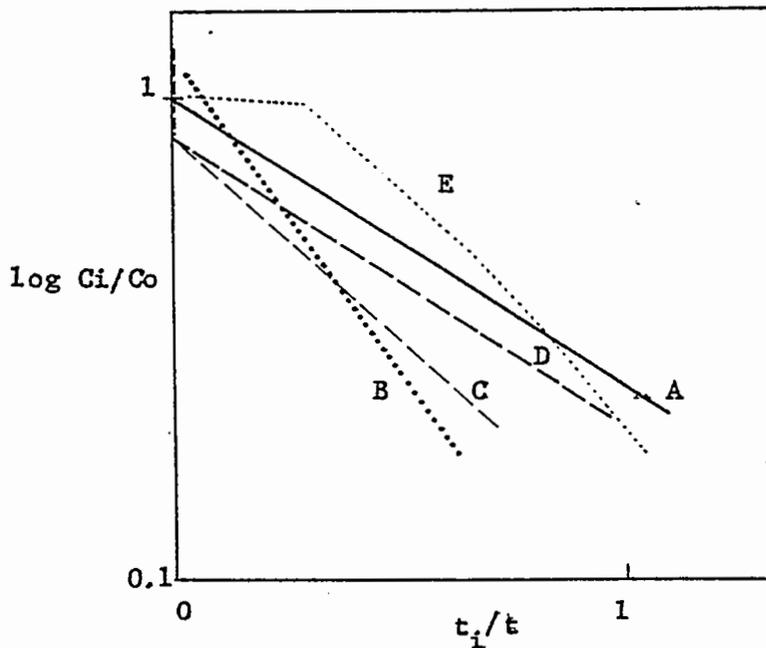
Figure 6 shows the typical washout curves obtained from a complete mixed reactor. In order to interpret the mixing data accurately, C_i/C_o and t_i/\bar{t} must be known. In real reactors like aeration tanks, it is extremely difficult to estimate C_o accurately due to turbulence and the volume occupied by air bubbles. Therefore, it is difficult to estimate the amount of short circuiting and dead space as indicated by the locations of the Y intercept. It appears that all lines in Figure 6 approach a straight line except line E, which represents a combination of plug flow and complete mixing.

Using a regression technique with the raw data, C_i and t_i , the washout rate was easily obtained and converted to the mean detention time \bar{t} , since the washout rate is equivalent to $1/\bar{t}$.

For the plug flow and dispersion models including the complete mixing plus plug flow system as represented by E in Figure 6, the regression technique could not be used to determine the mean detention time. The following formula was used (1):

$$\bar{t} = \frac{\sum_{i=1}^{i=\infty} C_i t_i}{\sum_{i=1}^{i=\infty} C_i} \quad (3)$$

The mean detention times estimated from the above equation for the pilot and 4 l units were not always identical to the values obtained by utilizing the regression technique. The discrepancy may be due to the fact that the wash-out curve is never completely measured and an infinite number of time intervals is not used in the final evaluation from the continuous curve. The results of both computational methods were summarized in Table II.



A : ideal complete mixing, B : complete mixing with dead space, C : complete mixing with short circuiting and dead space, D : complete mixing with short circuiting, and E : complete mixing with plug flow.

Figure 6. Comparative Shapes of Tracer Curves of Complete Mixing

The mixing tests for the nominal 4 hour run in the 4 l unit was conducted under an unsteady-state condition, i.e. while the microbes were growing. The washout rates obtained from the regression technique were considered to combine both washout rate and solids accumulation rate. The solids accumulation rate was approximately 0.25 g/hr. The rate of mean detention time change per g of solids accumulated was estimated to be 0.086 hr/g. This information allows the actual mean detention time changes to be computed. The values are 0.28 hr for the 1st day mixing test result (13 hr operation x 0.25 g/hr solids accumulation rate x 0.086 hr/g), 0.22 hr for the 2nd day, 0.15 hr for the 3rd day, and 0.26 hr for the 4th day results. Since these changes are small compared to the fluctuations of mean detention times shown in Figure 7 it was felt that they could be neglected.

Figure 7 shows the mean detention time variations along with the detention times computed based upon the drained volumes. The comparisons are favorable when one considers the possible variation in the mixing values caused by the time required to conduct the mixing test. The drainage method does not have this problem. This suggests that the drained volume technique is a useful method to determine the mean detention time. Also, the problem of chloride interference in the COD test (not reported in this study) is avoided.

Parameters Defining Mixing Regimes

For a plug flow regime, the dispersion number has been used to indicate the degree of dispersion. A dispersion number of zero represents ideal plug flow, and a dispersion number that approaches infinity indicates an ideal complete mixing regime. The value of 0.02 is indicative of a small amount of dispersion, 0.025 an intermediate dispersion, and 0.2 a large level of dispersion.

Other parameters used to define mixing conditions are the Morrill Index (3) and the variance. If the Morrill Index is equal to 1, the flow regime will be ideal plug flow. The system is considered to approach complete mixing as the Morrill Index increases. The variance ranges from zero to 1 when the flow regime changes from plug flow to complete mixing.

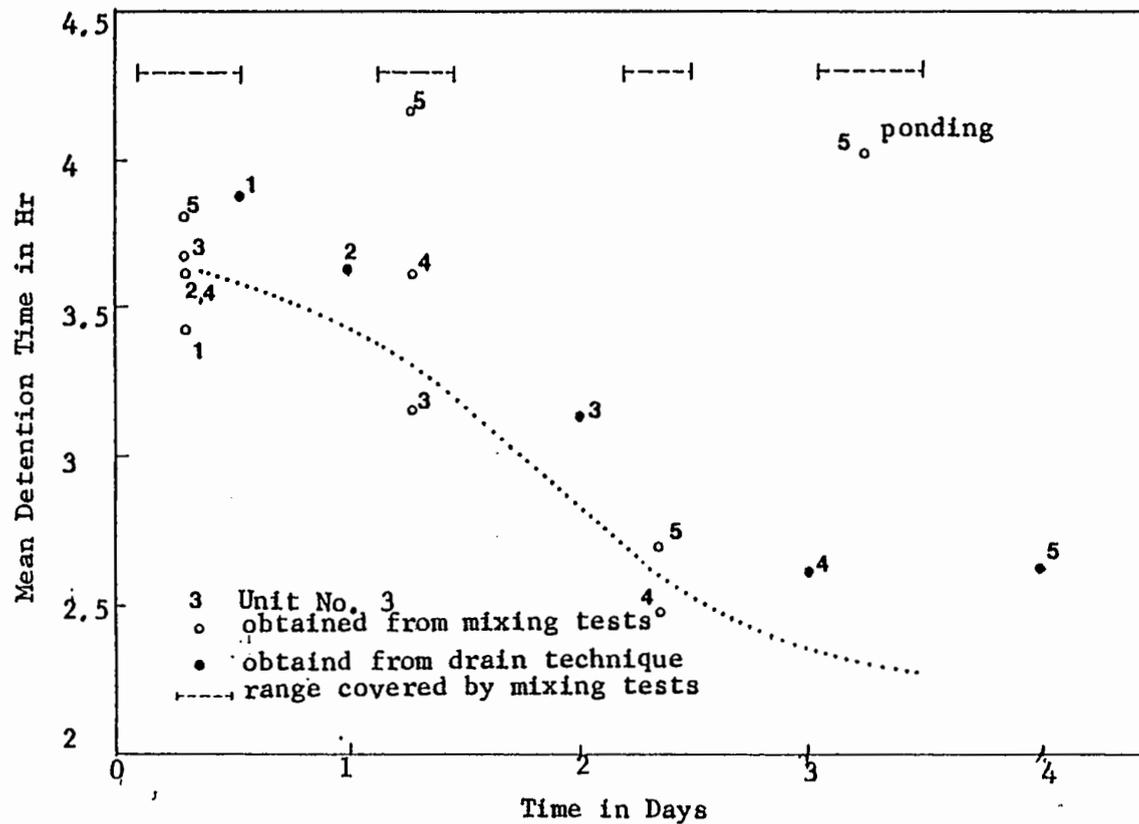


Figure 7. Mean Detention Time Variation by Microbial Growth-Nominal 4 Liter Unit

All parameters for defining the mixing characteristics have been developed for a specific reactor which is designed to achieve one of these ideal mixing conditions: complete mixing, plug flow, and dispersion models.

The following equation was used to obtain dispersion numbers for the fixed media reactor (1):

$$\sigma^2 = \frac{\sigma_t^2}{\bar{t}^2} = 2 \frac{D}{uL} - 2 \left(\frac{D}{uL} \right)^2 (1 - e^{-uL/D}) \quad (4)$$

where D/uL = dispersion number
 σ^2 = variance equals to σ_t^2/\bar{t}^2 , variance of a tracer curve
 σ_t^2 = variance in time units, hr²

Also, the variance, σ^2 , can be obtained from:

$$\sigma^2 = \frac{\sum t_i^2 C_i}{\sum C_i} - \bar{t}^2 \quad (5)$$

The time intervals for these equations must be the same regardless of the shape of the washout curve.

The variance data presented in Table V illustrates the sensitivity of equation 4 to actual tracer studies. Mixing data from the 4 hour, 4 l units with media and without microbial growth were used to compute σ^2 . The mean detention time and σ_t^2 values were determined by regression.

It appears that the σ_t^2 values were very sensitive to the time range and intervals. Consequently, σ^2 and the dispersion number became sensitive.

If the confidence limits of the regression lines are considered, σ^2 for unit No. 1 can be reduced to 1.28 and σ^2 for unit No. 5 can be increased to 0.93. The σ^2 for the other units are close to 1.0 which represents ideal complete mixing. The σ^2 values obtained from equation 5 and reported in Table II are in agreement with those computed and shown in Table V. Both methods confirm that an ideal complete mixing condition existed. However, these results do not agree

with the mean detention time evaluations shown in Table II. The theoretical values do not agree with the moments values of equation 5 which indicates that dead spaces existed in the completely mixed reactors.

Table V Variance of Mixing Tests

Unit Nos.	\bar{t} (min)	Time Range in Data (min)	Number of Intervals	σ_t^2 (min ²)	$\sigma^2 = \frac{\sigma_t^2}{\bar{t}^2}$
1	258	2 - 1020	22	94818.88	1.42
2	248	1 - 640	20	51745.99	0.84
3	238	1 - 625	18	50802.26	0.90
4	219	1 - 610	16	44130.73	0.92
5	229	1 - 600	16	43997.58	0.75

Effective Volumes of Media System

The mixing curves generally indicate that the test units were hydraulically completely mixed with dead spaces. Tolaney's (4) results, which showed that solids growth on the media was uniform with depth, indicated that his unit was biologically completely mixed in terms of McKinney's viewpoint (5). The circulation caused by the aerator continued to expose the substrate to the microbes and at the same time produced a relative velocity between the microbes and the food which would enhance the process kinetics (8).

The fixed media biological system is considered to be a combination of two different hydraulic zones; an effective zone which can be defined as the volume where food and oxygen are available to the microbes and an ineffective zone near the solids-media interface where substrate and oxygen supplies are limited (see Figure 8). The effective zone is believed to follow normal activated sludge concepts, while the latter zone facilitates thickening and digestion processes. In this study the effective zone is assumed to be the same as the effective hydraulic volume and the ineffective volume is assumed to be the same as the hydraulic dead space measured by the mixing tests.

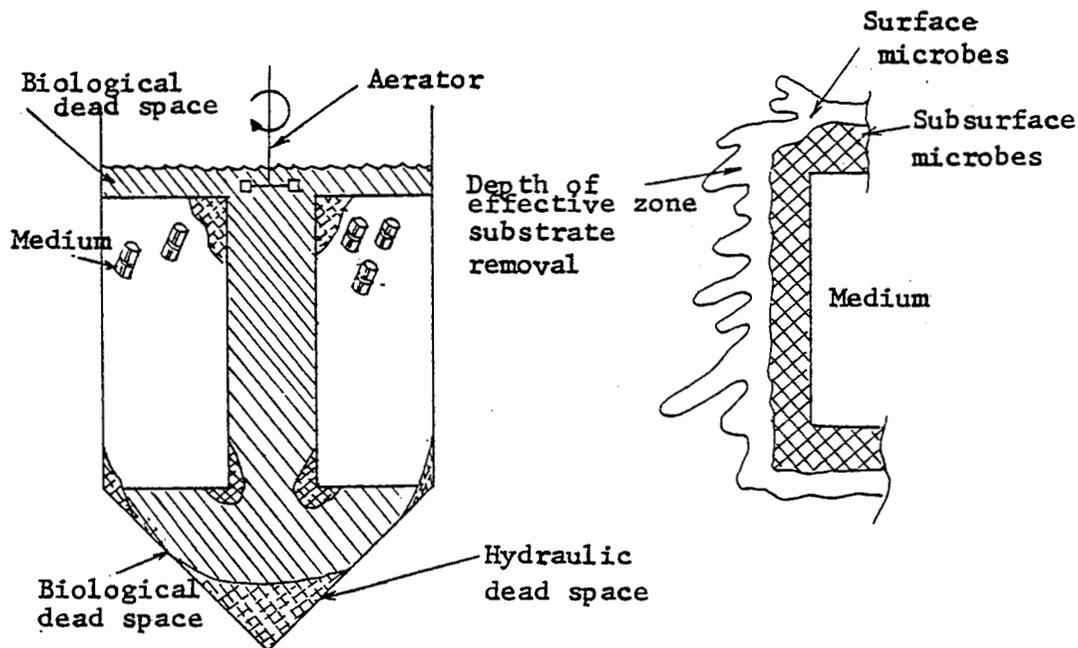


Figure 8. Illustrative Diagram for Hydraulic and Biological Effective Spaces

The biological effective volume of the fixed media process will be less than its hydraulic effective volume because the microbes are not distributed throughout the latter volume. These volumes include the draft tube, hopper, and free board liquid. The biological effective volume provides the detention time available for substrate removal. For the 4 l unit, the biological dead space was 0.8 l.

The average liquid volume drained for the 16.5 l unit was 14 l for the 1.6 cm media, 15 l for the 3.8 cm media, and 17 l for the 8.9 cm media operation. If the draft tube and bottom hopper volumes are subtracted from these volumes, the actual effective liquid volumes would be about 11.5 l for the 1.6 and 3.8 cm media units, and 12.5 l for the 8.9 cm media.

The operating biological effective volume of the pilot unit was computed to be about 11 cu m which is about 74 percent of the total volume. For endogeneous respiration, the biological effective volume would be the volume of the media

zone or 12.2 cu m .

The hydraulic flow patterns in the fixed media system was considered to be more vigorous in the upper portion of the media zone due to the mixing and turbulence of the aerator and air bubbles. The air bubbles tended to go upward, while the liquid was forced to go down. The microbial flocs attached to the media were observed to vibrate with the flow. The flow was no doubt in the downward direction on a microscopic scale. This flow condition indicated microscopic mixing within the system.

A Proposed Design Application

As solids sloughing is a phenomenon that is experienced in a trickling filter, it appeared that minor sloughing always occurred in the fixed media units. This continuous discharge of solids is a form of solids washing which is associated with the hydraulic and biological forces. There exists a balance between these forces which permits the release of solids. However, the effluent solids level appeared to be mainly related to the food and microbial level in the system at the normal operating range (6).

Figure 9 shows the effluent VSS or TSS versus organic loading rates. The results of the nominal 4 l unit were plotted as VSS and the other tests were plotted as TSS. The VSS/TSS ratios ranged from 0.65 to 0.85 for the 16.5 l unit, and from 0.65 to 0.95 (average 0.70) for the pilot unit. Since the pilot unit was not equipped with an efficient solid wasting capability, some of the values show very high TSS concentrations in the effluent. All of the open data points could be eliminated if proper solids wasting could have been practiced. The open circles for the nominal 4 l unit operation were thought to be caused by the switch from a laboratory aeration-only to a fixed media operation, and by operation above the maximum solids holding capacity. The lowest line indicates the expected minimum solids level in the effluent of a 1.6 cm media system. It appeared that the effluent solids level increased with the increasing size of media, and that the solids variations were greater with the bigger media. Stackley (7) and Tolaneys' (4) data were plotted without detention time corrections.

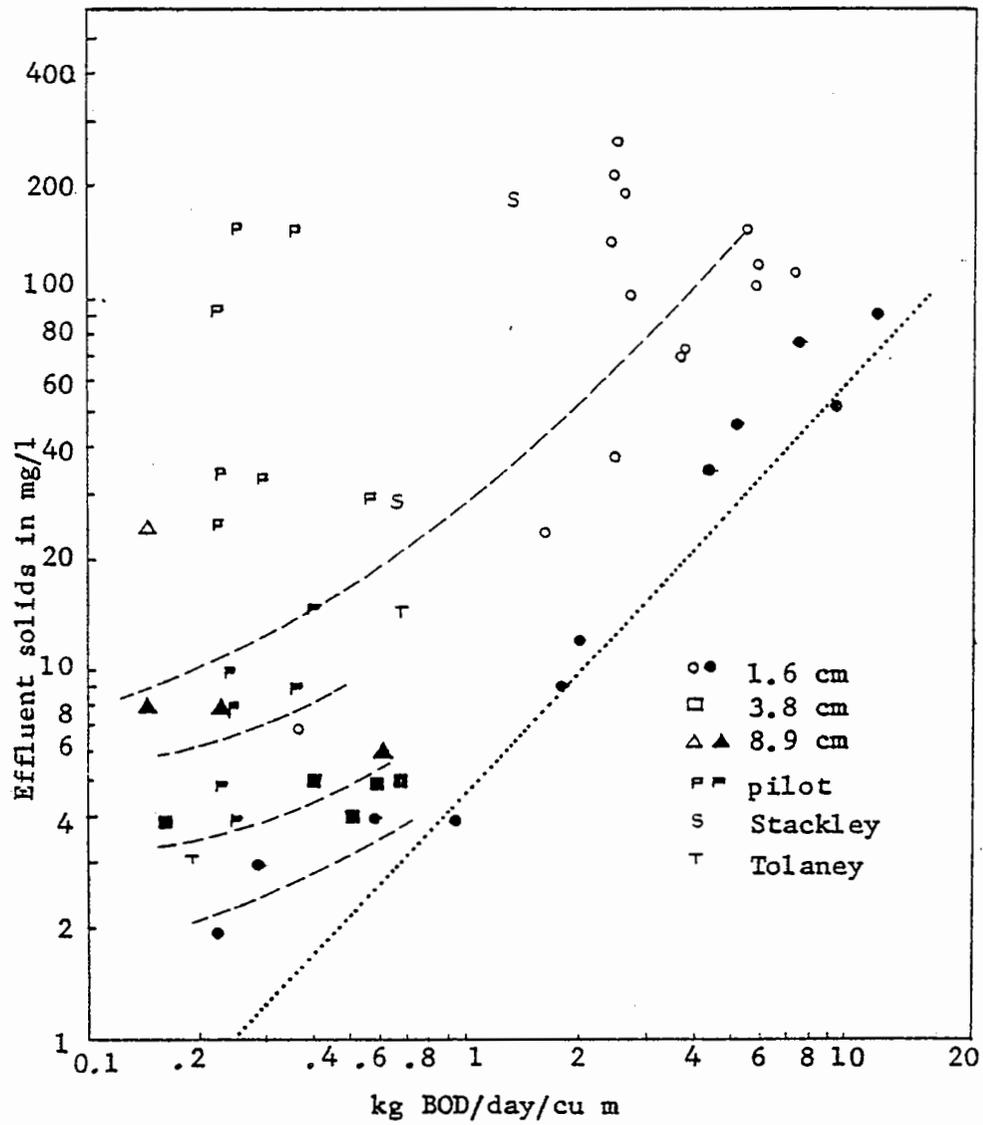


Figure 9. BOD Loading vs Effluent Solids

Since the fixed media system of this study generally approached a complete mixing flow pattern, the design of such a system can be made by using Figure 9 and McKinney's design equations (5). However, the hydraulic and solids retention times vary with time in the fixed media biological system, and three different detention times should be used: the biological detention time (\bar{t}^*), the nominal detention time (t), and the detention time of the media zone (t_1). For a better understanding of the design procedure a hypothetical fixed media plant treating domestic wastewater of 3,785 cu m/day with an influent BOD of 234 mg/l and SS of 192 mg/l is given as follows.

Assuming that the treatment requirements are 90 percent removal of BOD and TSS, the loading rate can be obtained from Figure 9 (effluent TSS = 19 mg/l). The maximum loading rate for this treatment plant at its maximum efficiency is about 4 kg BOD/day/cu m and 0.6 kg BOD/day/cu m at the minimum efficiency. An average loading rate of 1.6 kg BOD/day/cu m is selected.

The required volume of reactor can be computed as:

$$\begin{aligned} \text{total kg BOD applied/day} &= 885.7 \text{ kg} \\ \text{volume of media of reactor to be required based} \\ \text{on BOD loading} &= \frac{885.7}{1.6} = 553.5 \text{ cu m} \end{aligned}$$

Assuming 8.9 cm flexirings are utilized, the maximum solids holding capacity would be 64.6 g/sq m (see Table IV). The hydraulic dead space can be computed assuming a 6 percent solids accumulation:

$$\begin{aligned} &\frac{(\text{volume of media})(\text{wt. of total solids accumulated})}{\% \text{ solids}} \\ &\times \text{specific surface area of media} \\ &= \frac{55.35(\text{cu m}) \times 64.6 (\text{g/sq m}) \times 92.1 (\text{sq m/cu m})}{0.06 \times 1000 (\text{g/kg}) \times 1000 (\text{kg/cu m})} \\ &= 59.3 \text{ cu m} \end{aligned}$$

$$\begin{aligned} \text{Adding 7 percent loss in volume due to media, the re-} \\ \text{quired effective volume} &= 553.5 + 54.9 + 0.07 \times 553.5 \\ &= 647.1 \text{ cu m} \end{aligned}$$

Assuming that the biological dead space is about 20 percent of the total reactor volume, then the total reactor volume will be about 776 cu m without the media.

Now, the detention times which will be used for the computation are: $\bar{t}^* = 3.5$, $t_1 = 3.9$ and $t = 4.7$ hrs.

Assuming a metabolism constant K_m [from equation $F = F_i / (K_m \bar{t}^* + 1)$ where $F =$ effluent soluble BOD, $F_i =$ influent BOD] of 13.3/hr at the beginning of the operation and 4.5/hr at the end of operation and using the biological effective time, \bar{t}^* , the effluent soluble BOD can be computed to be 4.9 and 14 mg/l.

The maximum active mass M_a in the system [from equation $M_a = K_s F / (K_e + 1/t_s)$, where $K_s =$ synthesis constant, and $K_e =$ endogenous respiration rate] can be obtained from $K_s = 2.6$ /hr, and $K_e = 0.03$ /hr with continuous operation. The solids retention time t_s is long enough to assume $1/t_s = 0$ for the computation:

$$M_a = \frac{14 \times 2.6}{0.03} = 1,213 \text{ mg/l}$$

The maximum solids holding capacity is equivalent to:

$$M_t = \frac{64.6(\text{g/sq m}) \times 92.1(\text{sq m/cu m}) \times 1000(\text{mg/g})}{1000(\text{l/cu m})}$$

$$= 5,950 \text{ mg/l}$$

The effluent BOD values are computed using the effluent TSS data shown in Figure 9:

$$\text{From Total eff. BOD} = F + 0.8 M_{a_{\text{eff}}}$$

$$\text{where } M_{a_{\text{eff}}} = M_{t_{\text{eff}}} \times \frac{M_a}{M_t}$$

Therefore,

$$\text{Total Eff. BOD (minimum)} = 4.9 + 8 \times 0.8 \times \frac{1,213}{5,950} = 6 \text{ mg/l}$$

$$\text{Total Eff. BOD (maximum)} = 14 + 40 \times 0.8 \times \frac{1,213}{5,950} = 21 \text{ mg/l}$$

The carbonaceous oxygen requirement dO/dt can also be computed by using the equation $dO/dt = 1.5 (F_i - F)/t - 1.42 (M_i + M_e)/t$, where M_e = endogenous mass. Since $(M_i + M_e)/t$ is difficult to obtain and essentially the system is designed for solids accumulation, the oxygen requirement can be computed as:

$$dO/dt = 1.5 \times 234/4.7 = 74.7 \text{ mg/l/hr}$$

The average daily solid production was estimated to be about 530 kg/day by utilizing an empirical design equation with t_1 developed during the course of this study (6), and this means that within about 6 days from the previous solids wasting, the system would again approach the maximum solids holding capacity.

This design example was made on the basis of a maximum loading rate and a possible maximum oxygen transfer rate. Actually, a practical design must be made on the basis of the ease of operation and maintenance with a proper consideration for solids management. This may mean increasing the reactor size to accommodate a particular design situation.

SUMMARY AND CONCLUSION

The fixed media biological process can be designed to approach a hydraulically complete mixed system, but the flow pattern tended to approach a plug flow regime as the process approached its maximum solids holding capacity.

The fixed media process is considered to be a combination of both hydraulic and biological effective and ineffective volumes or spaces. The biological volume for substrate removal will be less than its hydraulic volume, because the microbes are not distributed throughout the latter volume. The draft tube, hopper, and free board liquid volume is the biological dead volume. The hydraulic dead space like the solid-media interface can be biological effective space for endogenous respiration. The drainage volume of the media process represents a hydraulic effective volume. Comparisons of both mixing and drain methods for computing mean detention times are favorable when one considers the possible variation in the mixing values caused by the time required to conduct this evaluation.

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THE EFFECTS OF HYDRAULIC VARIATION ON
FIXED FILM REACTOR PERFORMANCE

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INTRODUCTION

Rather than describe the performance of fixed film reactors in general; the objective of this paper is to describe the response of two attached growth systems, viz. rotating biological contactor (RBC) and trickling filter (TF) to hydraulic variation. For each system, a mathematical model was selected based on a review of available models and theoretical analysis of the unit processes involved. The model parameters were then "bounded," i.e., the range of probable values that a parameter could take were defined. The models were then arranged with effluent quality (in these cases, soluble BOD concentration) as the dependent variable and influent flow rate as the principal independent variable. The models were then "exercised" over the range of probable parameter values and the output, i.e., the variation in effluent quality, was graphically displayed as a function of flow rate and any other independent variables used in the model. Where practicable, model output with unsteady flow was divided by model output with steady flow to reduce the

effect of uncertainty in model parametric values used in the simulations. Also, the models were used to simulate the 24-hour average effluent quality that would be expected as a function of diurnal hydraulic variation.

AVAILABLE MODELS

In order to analyze the effect of peak flows on fixed film reactors, it is necessary to analyze, in some detail, the physical and biochemical processes that are taking place.

In a suspended growth system, maintenance of environmental conditions, especially with regard to dissolved oxygen tension, can be controlled independently of other process variables through the use of mechanical aeration. In attached growth processes, maintenance of oxygen tension, in the vicinity of the biological reaction, cannot be accomplished independently. That is, the mechanical actions that affect the dissolved oxygen level also affect the physical transport of the reaction components and the biochemical reaction rate itself. Another significant difference between fixed film and suspended growth systems is that the transport of the reactants in a suspended growth system may be considered to be well described for ideal types of reactors, especially complete mix reactors. Therefore, by having the ability to independently control oxygen tension and the physical transport of materials to the reaction site by mechanical means, the only significant variable to be analyzed in suspended growth systems is the biochemical reaction rate. In attached growth systems, however, changes in flow rate or influent substrate concentration may cause significant changes in substrate transport, substrate reaction and the effective dissolved oxygen level. Because of the interrelationship of these phenomena, a detailed analysis of attached growth reactors is required in order to predict the effects of diurnal flow variation on performance.

A fundamental assumption used in analysis of attached growth systems is that, although hydraulic variation may affect the instantaneous rate at which filter humus is removed from the reactor, the average rate of humus loss is independent of hydraulic variation, for the same average flow. (A good assumption for laminar flow.) In light of

this assumption, the objective of this paper is to characterize the variation in effluent soluble BOD concentration (SBOD) caused by hydraulic variation. In order to construct a graphical display of this response, it was necessary, as described above, to carefully analyze the various physical and biochemical processes in an attached growth reactor.

The Effect of Hydraulic Variation on RBC Soluble BOD (SBOD) Concentration

The analysis of RBC performance with respect to SBOD removal has occupied many theorists and practitioners. For domestic sewage there are at least four phenomena that must be described and interrelated in order to describe RBC performance: electron donor (substrate) transport and kinetics, and electron acceptor (DO) transport and kinetics. If we assume domestic sewage to be adequately characterized by SBOD, then the reaction in any one stage of an RBC train is that between SBOD and DO. For each of those components of reaction, there can be constructed a transport model and a kinetic model. There are, therefore, at least four expressions that must be developed, one for each component described above. In the overall model, these expressions must be linked together, as only one of the components will be controlling the overall observed rate of reaction at any time.

There are two relatively recent theoretical approaches that best illustrate the difficulties of constructing a mechanistically sound model for attached growth processes. The approach by Williamson and McCarty (1) results in an algorithmic solution which balances the surface flux of substrate into the biofilm with the surface flux of substrate that can be assimilated or oxidized within the biofilm. Williamson and McCarty's model requires specification of seven coefficients: k , K_S , D_w , D_c , X_C , L_1 , and L_2 (see Notation for explanation). While methods were presented for measuring, calculating or estimating these coefficients, the predicted performance is sensitive to the set of coefficients selected. Unfortunately, many of the coefficients must be estimated as there is no simple method for direct evaluation. Williamson and McCarty's work, however, does clearly show the effect of the various components of the system on

overall performance and draws explicit distinctions between transport of the electron donor and the electron acceptor, and the reaction rate control by the electron donor or the electron acceptor. While the model has been verified for a specialized experimental device, the model has not been directly applied to RBC or trickling filters, nor has the variation of the model parameters been characterized as a function of the geometry of the system and the influent flow variation.

Among the most detailed analysis to date (for RBC systems) is that by Grady and Lim (2). Their analysis, which relied heavily on fundamental analysis of hydrodynamic transport, requires specification of some 17 parameters, or coefficients, which allow the analyst to determine which of the four phenomena described above are controlling the overall rate of reaction for any particular condition. Based on this analysis, a figure similar to Figure 1 may be developed which shows effluent substrate concentration as an explicit function of influent flow rate and influent SBOD concentration and an implicit function of the other 15 parameters which may affect performance. It is worth noting that many of these implicit parameters depend upon the geometry of the RBC system, the depth of media submergence and the angular velocity of the RBC. While Grady and Lim's analysis could be applied to trickling filters, it will be shown herein that the more general approach by Harremoes is directly applicable to trickling filter systems.

While predictions such as those shown in Figure 1 are very helpful for a single RBC unit, they provide little help in predicting what range of performance might be expected for RBC systems. A notable result, however, of Grady and Lim's analysis is that the apparently superficial analysis described by Clark et al. is in fact a reasonable basis for characterizing or correlating data for a particular system. Clark et al. (3) derived a Monod-like expression (Equation 1 shown on Figure 2) by writing a material balance around an RBC stage which relates the removal of substrate per unit area (R) to the maximum areal rate of removal (P), a pseudo-half velocity coefficient (K) and the equilibrium concentration of SBOD in that stage (S). Equation 1 may be linearized by inverting (Equation 2 - Lineweaver-Burk) or by multiplying the inverted expression by the product of P and R (Equation 3 - Eadie-Hofstee). Figures 2, 3, and 4 show a plot of

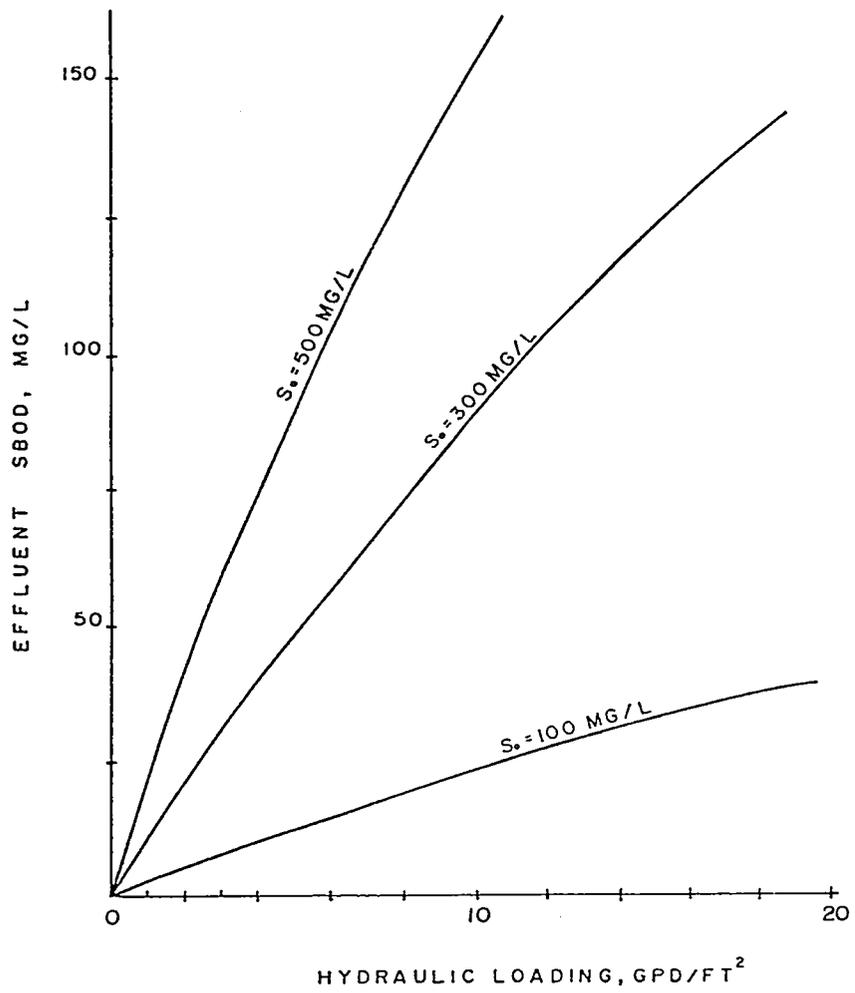


FIGURE 1: RBC Performance (After Grady and Lim)

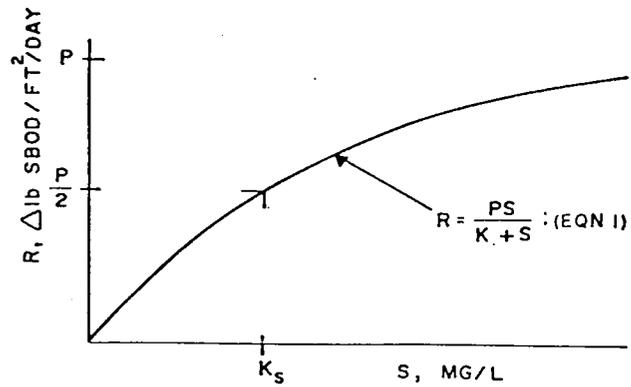


Figure 2. RBC Data Correlation (After Clark et al).

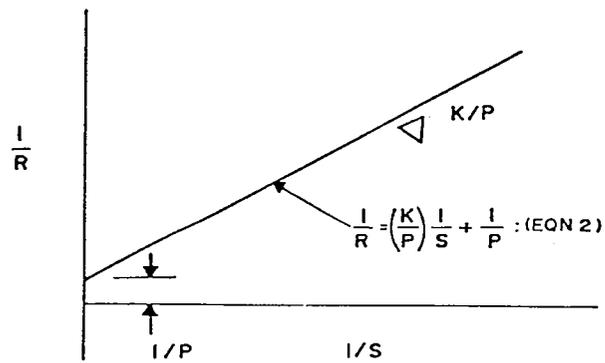


Figure 3. Lineweaver - Burk Linearization of Equation 1.

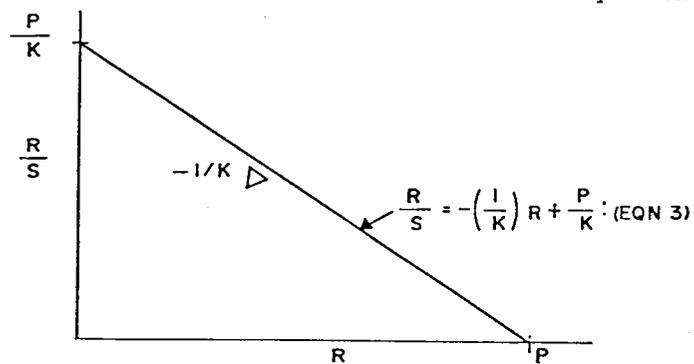


Figure 4. Eadie-Hofstee Linearization of Equation 1.

Equations 1, 2, and 3, respectively. In the absence of any corroboration, however, it would appear hazardous to assume that the expression proposed by Clark et al. would be valid over a wide range of BOD concentrations.

The degree of hazard is related to the fact that one of at least four processes may be controlling the overall rate of reaction. It would appear probable that as the controlling process changes over the range of flow or substrate concentration, the parameters P and K defined by Clark et al. in Equation 1 might also vary, thereby changing the expected performance significantly.

Some verification of Equation 2 (Figure 3) can be obtained, however, by taking data from a figure similar to Figure 1 (developed by Grady and Lim) and replotting on the RS plane (similar to Figure 2). It will be seen that even though Grady and Lim's model is intended to accommodate dominance by any of the four processes described before, it plots as a smooth curve on the RS plane. Therefore, for an RBC system which consists of similar geometry stages, the only variables in the Grady and Lim model which could significantly change the RS plot are the kinetic parameters which are themselves related to the composition of the substrate only. The benefit of using the RS plane plot is that the 17 parameters described by Grady and Lim are reduced to two: P and K.

Experience in multi-stage activated sludge systems (including lagoons) has indicated that within a complete mixed unit, the biochemical reaction may be well described by a single set of kinetic parameters which describes the reaction rate in equilibrium with a soluble substrate level. However, when the effluent from a single mixed unit becomes the influent to a second reactor, the second reactor will exhibit a markedly different reaction rate. This will reflect not only the change in the soluble substrate concentration, but also changes in the fundamental reaction rate parameters, the maximum rate of substrate removal, and the half velocity coefficient. Therefore, although Clark et al. have indicated a method to correlate RBC stage data which appears to be rational from a fundamental point of view, as shown by the work of Grady and Lim, it is apparent that more than one curve on the RS plane may be required to describe an RBC system. In effect, one curve per stage may be required to provide adequate characterization.

If stage curves can be developed for an RBC system on the RS plane, this type of plot has many advantages in

terms of predicting effluent quality. In Figure 5, it can be seen that if a line is started at coordinates $(S_0, 0)$ and extended with a slope of $-(Q/A)$, where A is the stage surface area, the abscissa at which it intersects the stage line will be the effluent quality for that stage. If a line with a slope equal to $-(Q/A)$ is extended from that abscissa to intersect the next stage line, the abscissa at which it intersects the second stage line will be the effluent quality from the second stage and so forth. For any stage, therefore, the effluent quality is a function of P , K , S_0 and (Q/A) and can be expressed mathematically (Equation 4) as well as graphically.

Interstage data were collected to determine if RBC data could be shown to fall within definite areas on the RS plane. Based on a review of available interstage data, first and second stage performance was found to fall within the upper envelope shown in Figure 6, and third and fourth stage performance were found to fall within the lower envelope in that same figure. These envelopes correspond to the variation in P and K values shown.

Equation 4 was derived to calculate the fractional BOD remaining from the i th stage as a function of P , K , S_{i-1} , and (Q/A) . The data were plotted in envelopes where one envelope represented the probable range of P , K , and S_{i-1} for the first and second stages, and the second envelope represented the probable variation of P , K , and S_{i-1} for the third and fourth stages. The results, shown in Figure 7, are a plot which shows the fractional SBOD remaining as an explicit function of (Q/A) , with the values of SBOD for any (Q/A) falling within the stage envelopes as long as the values of P , K , and S_{i-1} are within the ranges shown. For any number of stages, the final effluent quality will be equal to the influent quality times the product of the fractional SBOD remaining as a function of the overflow rate (Equation 5).

As the time scale for transport of substrate into the attached growth film is small compared to the time scale of hydraulic variations (except for pumped systems), it may be assumed, as a conservative case, that hydraulic variations will produce an instantaneous effect in terms of change of effluent quality. This would represent the maximum probable effect of hydraulic variation on RBC SBOD quality.

Figure 7 was used to construct an effluent quality profile for a four stage RBC system. Two different staging sequences with the same total area were assumed.

FOR ANY STAGE i : $S_i = \frac{1}{2} \left[(B^2 + 4S_{i-1} \cdot K)^{\frac{1}{2}} - B \right]$ (EQN. 4)

WHERE : $B = \left[P \left(\frac{Q}{A} \right) + K - S_{i-1} \right]$

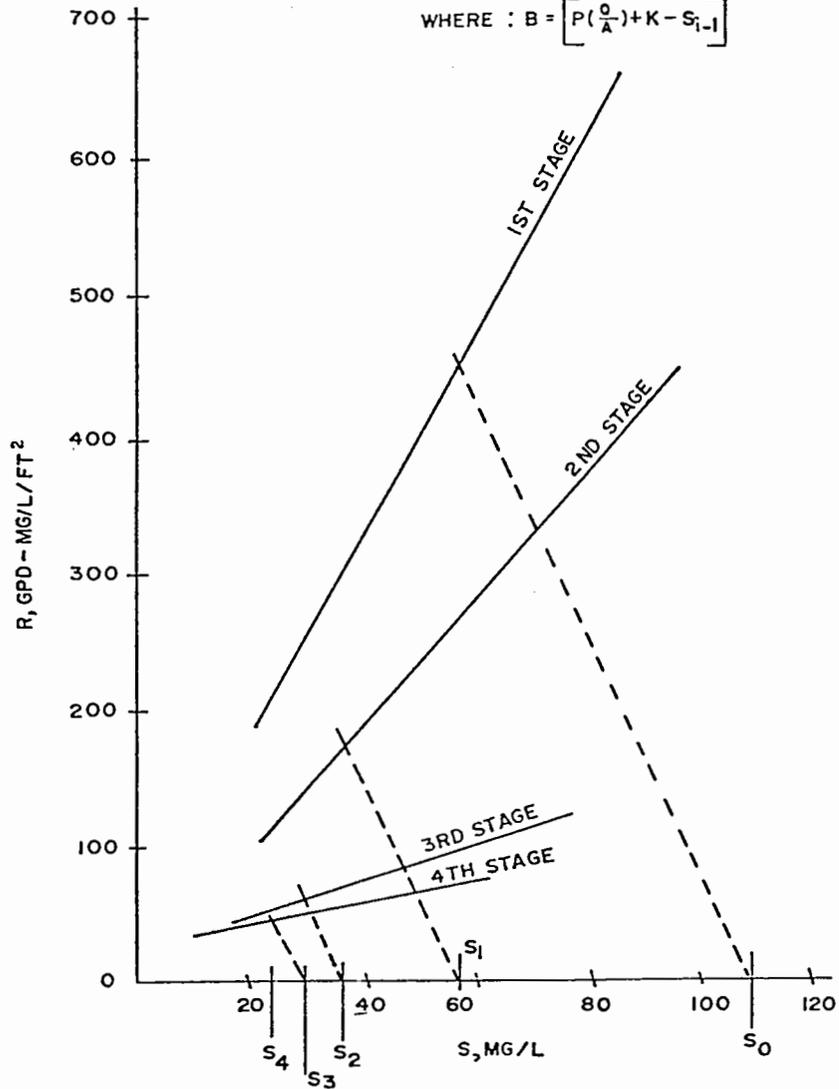


Figure 5. RBC Performance Calculation

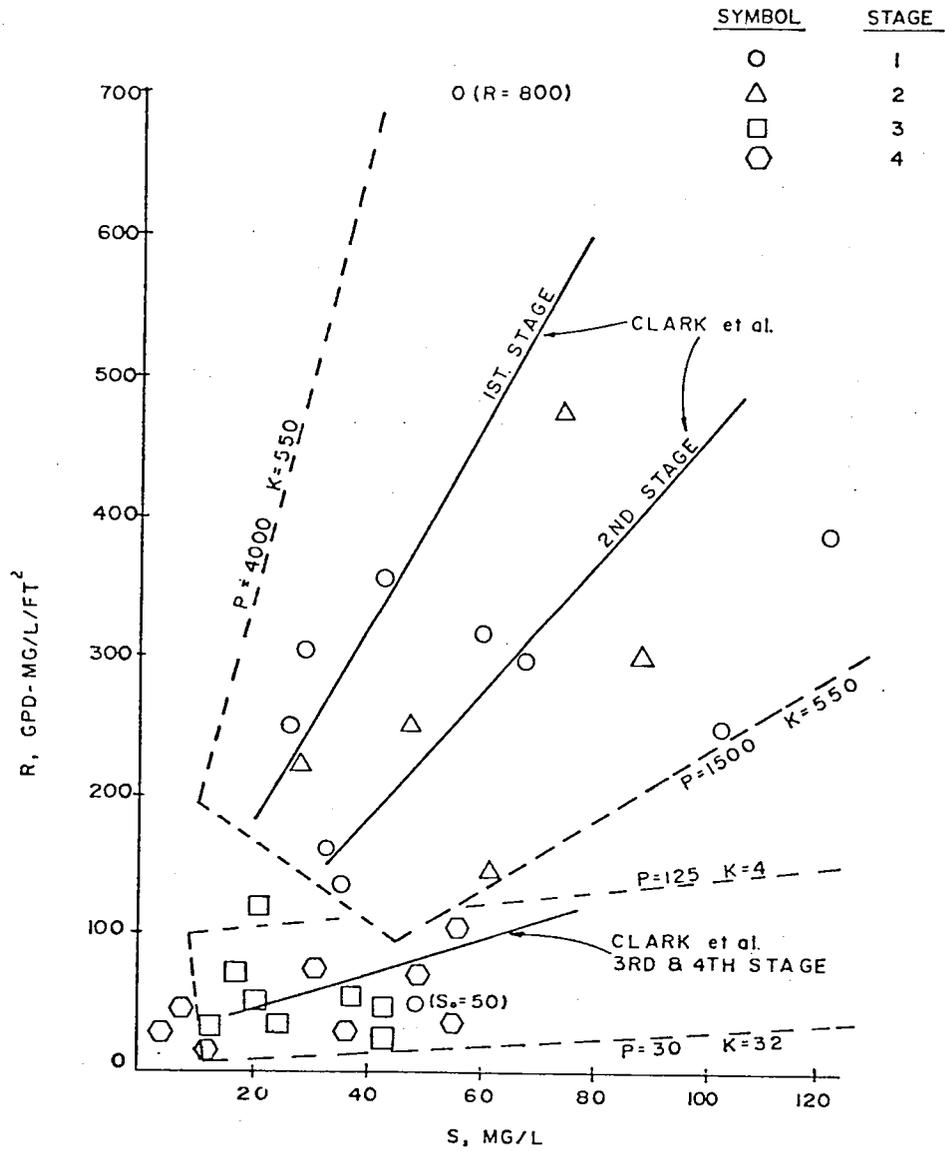


Figure 6. RBC Interstage Data

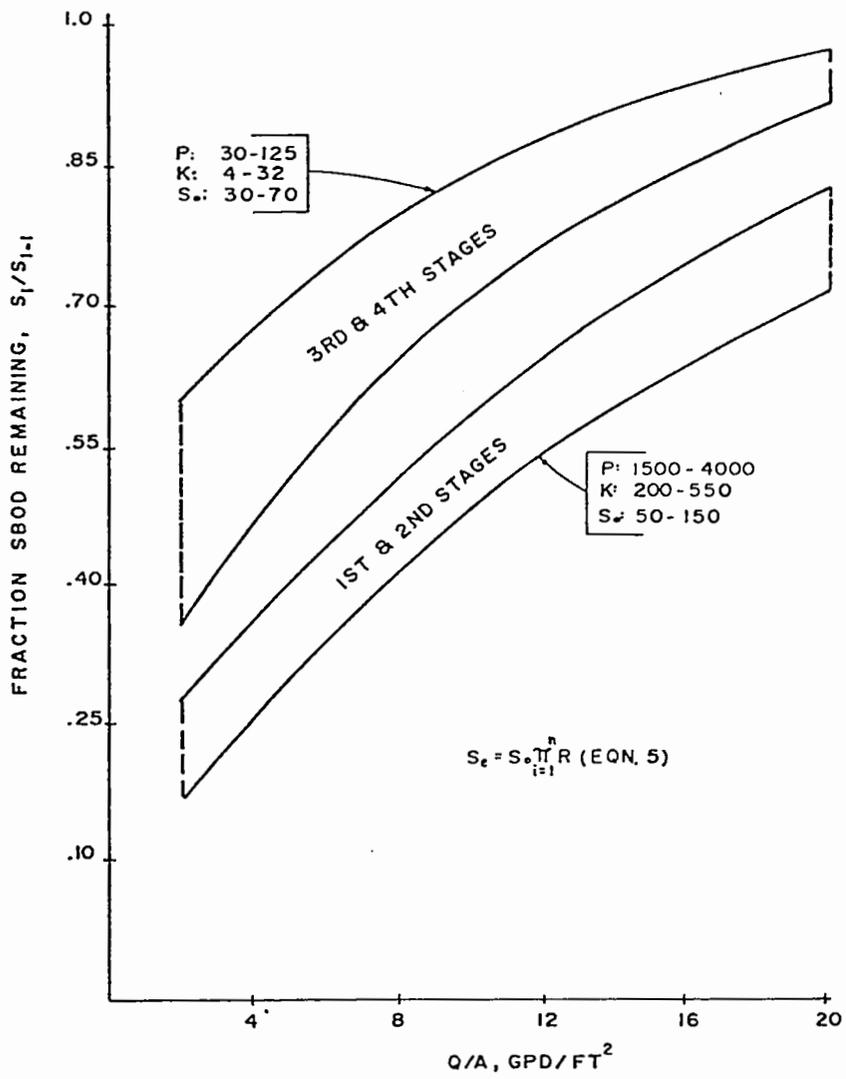


Figure 7. RBC Stage Performance

One system consists of two parallel trains, each with one unit per stage with a unit area equal to 100,000 square feet. In the other system, the eight units are distributed in four stages in the ratio 4:2:1:1. Figure 8 shows the assumed configurations and the corresponding (Q/A) for each stage.

The RBC system simulations were conducted by using Figure 7 to predict effluent quality as a function of instantaneous areal flow rate (Q/A) for the two systems shown in Figure 8. Based on the development of Figure 7, it was assumed that the range of system performance was well described by the upper and lower boundaries of the appropriate state envelopes shown in Figure 7. Therefore, the only model independent parameters were:

1. (Q/A) as a function of time.
2. RBC system configuration--4211 or 2 @ 1111.
3. Kinetic activity--poor (the upper boundary of each stage envelope in Figure 7) or good (the lower boundary of each stage envelope in Figure 7).

The parameter (Q/A), as a function of time, was defined to be of three possible types:

1. Steady flow; i.e., (Q/A) is not a function of time.
2. "Normal" Publicly Owned Treatment Works (POTW) diurnal flow.
3. An arbitrary flow variation representing equalized industrial flow.

Type 2, "Normal" POTW, diurnal flow was defined from a survey of over 100 POTW (4). The survey involved collecting 20 days of flow records (10 days in spring and 10 days in summer), along with the following sewer system data: Average system age; Groundwater variability; Soil types; System geometry; Amount of pumping in the sewer system; and, Average diurnal flow. The survey results indicated that two distinct types of influent flow variation could be distinguished. The first type was defined as normal diurnal flow variation or "normal," and the second type was defined as pump dominated diurnal flow variation or "pump dominated." The "normal" type was found to be independent of all of the sewer system characteristics described above except for pumping at the POTW headworks, even though the system average ages ranged

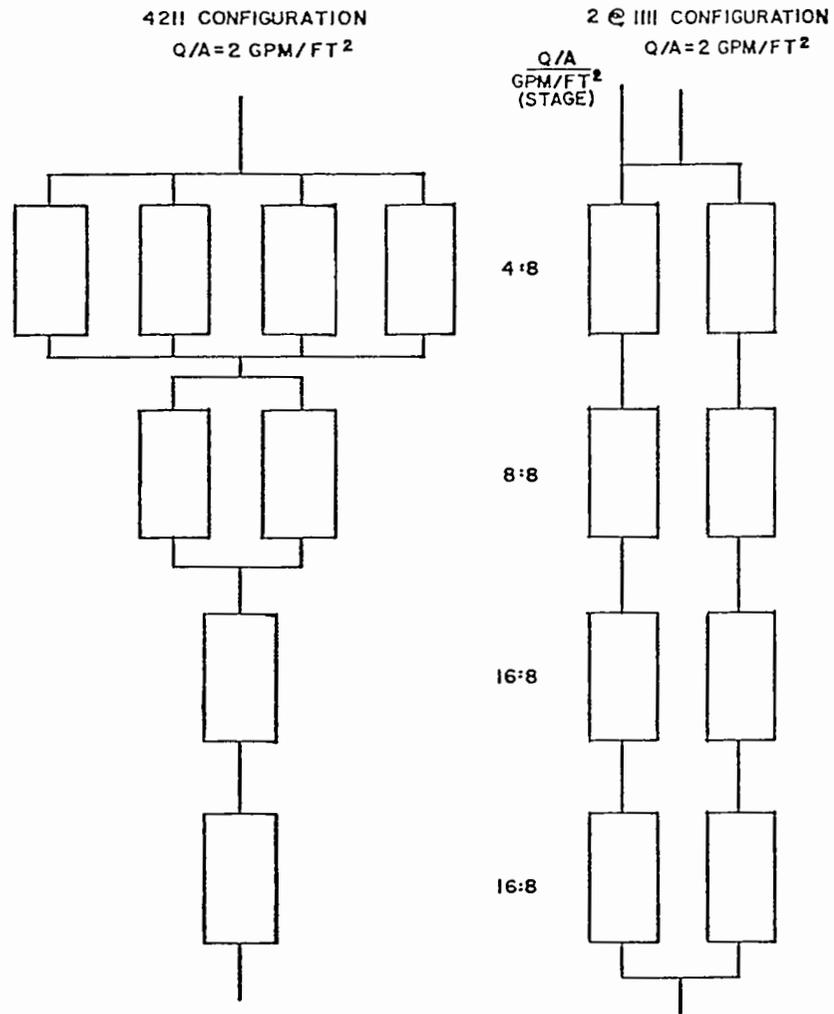


Figure 8. RBC Stage Configurations Used for Simulation

from 10 to 100 years, the groundwater variability was such that some systems were always immersed and others never immersed, the soil types ranged from clay to sand, the system geometry ranged from globular or clustered to extremely elongated, and the average annual flow ranged from 0.5 to 70 million gallons per day (mgd). The diurnal flow, in summary, for systems that were not pumped at the POTW headworks, was found to be relatively similar.

For modelling purposes, the "normal flow" was defined to be the curve shown in Figure 9, where Q_0 is the average daily flow for any given day for the POTW. The ordinates are, in effect, a plot of diurnal peaking factors, with the maximum ordinate being the Maximum Hourly Peaking Factor (MHPF). Type 2 flow ("normal flow"), therefore, was defined by the curve in Figure 9 for a given MHPF. This was done by exaggerating the deviations of the curve in Figure 9 from the daily average flow rate by the ratio of the MHPF to 1.23, the MHPF in Figure 9. This procedure is shown arithmetically in Equation 6 and graphically in Figure 10.

The performance of the RBC system as a function of peaking was simulated by the procedure described below for "normal" flow. Table 1 indicates the fraction of time that the influent flow is within each flow interval as a function of the MHPF. For each RBC stage, Figure 7 was used to determine the probable range of effluent quality, i.e., for each overflow rate for each stage, both the bottom and top of the appropriate stage envelopes were used to predict the fraction of SBOD remaining. The output of this simulation is for four types of flow (i.e., four MHPF for normal flow) in two types of RBC systems with two levels of response per system. There are, therefore, a total of 16 cases that were simulated for "normal" flow. The results of the simulation are shown in Figure 11. Figure 11 shows the normalized fraction of SBOD remaining as a function of MHPF, system configuration and system kinetic activity for "normal" flow.

Figure 11 indicates that peaking may cause an increase in effluent SBOD of up to 30 percent for the 2 @ 1111 configuration with good kinetics (the lower boundary of the stage envelopes in Figure 7). For the poor kinetic systems (the upper boundary of the stage envelopes in Figure 7), no significant deterioration in effluent SBOD is predicted. Clearly, the use of the terms "good" and "poor" is misleading. It may be more realistic to consider the

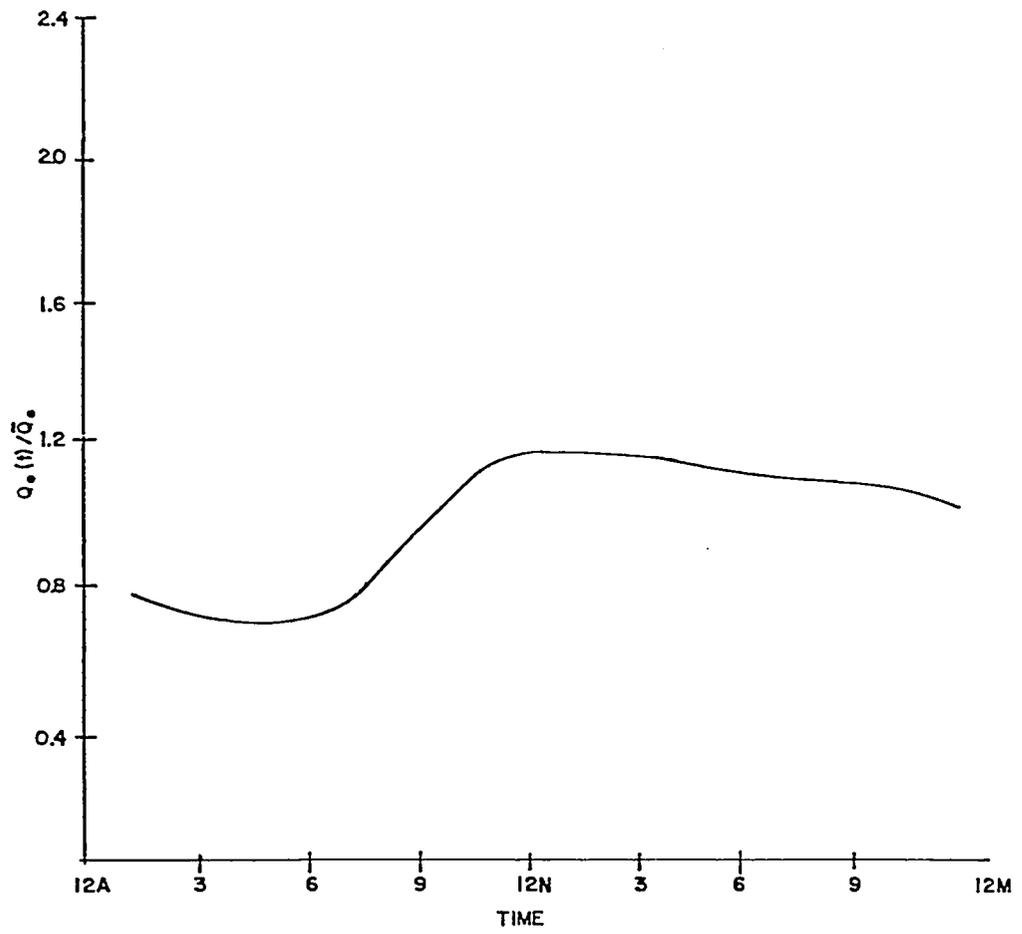


Figure 9 "Normal" Flow Variation

FOR ANY MAXIMUM HOURLY
FACTOR MHPF = X

$$\left[\frac{Q_o(t)}{Q_o} \right]_{\text{MHPF}=X} = \left[\frac{Q_o(t)}{Q_o} \right]_{\text{MHPF}=1.23}^{-1} \left[\frac{(X-1)}{(1.23-1)} \right] + 1 \quad (\text{EQN 6})$$

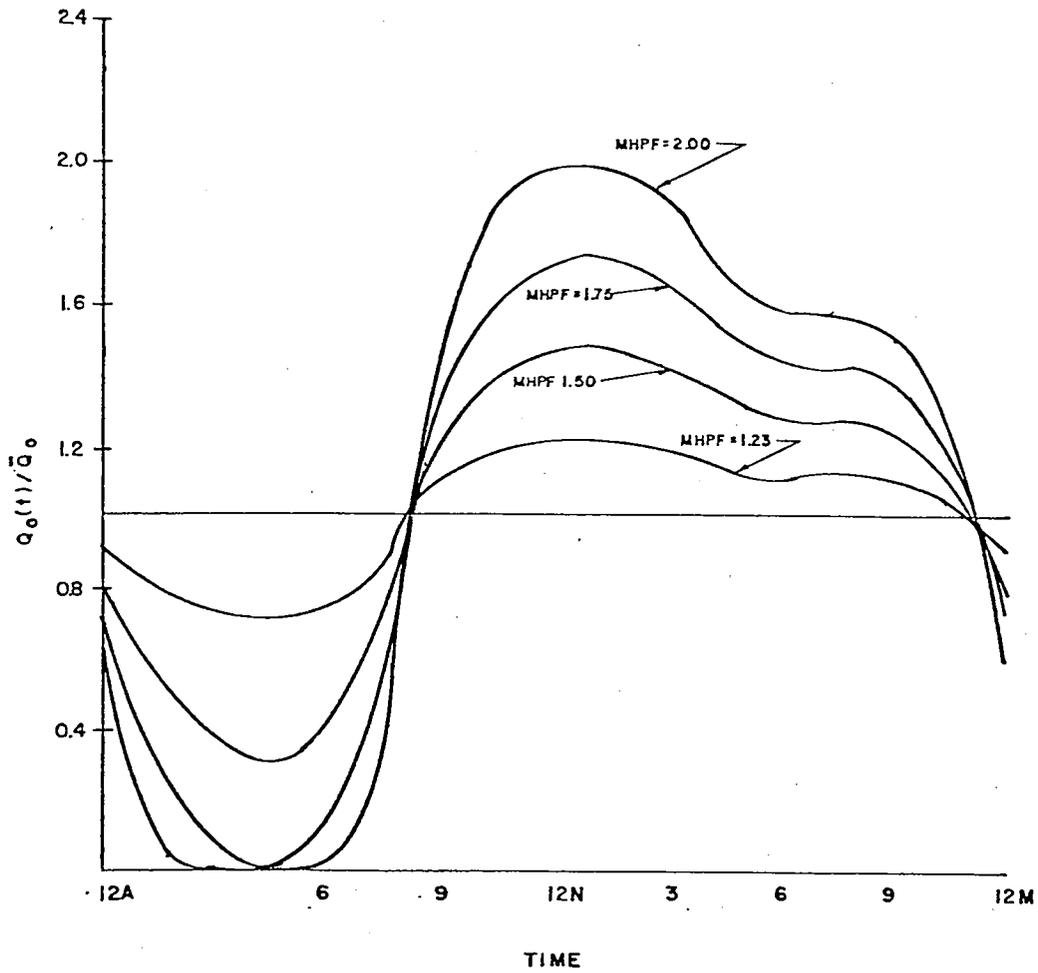


Figure 10 "Normal" Flow As A Function of MHPF

TABLE 1
CALCULATION OF RBC EFFLUENT QUALITY

$\frac{Q}{Q_0}(t)$	$f_{PF} = 1.23$	$f_{PF} = 1.5$	$f_{PF} = 1.75$	$f_{PF} = 2.0$
0 - .2			0.17	0.25
.2 - .4		0.17	0.13	0.04
.4 - .6		0.13	0.04	0.04
.6 - .8	0.29	0.04	0.04	
.8 - 1.0	0.21	0.08	0.04	0.04
1 - 1.2	0.50	0.04		0.04
1.2- 1.4		0.37	0.13	0.04
1.4- 1.6		0.17	0.28	0.25
1.6- 1.8			0.17	0.08
1.8- 2.0				0.22
0 - 2	1.0	1.0	1.0	1.0

CONF	KINETICS	$\frac{S_E}{S_0}$ (PF=1)	$\frac{S_E}{S_0}$ (PF=1.23)	$\frac{S_E}{S_0}$ (PF=1.5)	$\frac{S_E}{S_0}$ (PF=1.75)	$\frac{S_E}{S_0}$ (PF=2)
4211	Good	0.112	0.097	0.111	0.123	0.139
2@1111	Good	0.246	0.215	0.224	0.229	0.254
4211	Poor	0.125	0.124	0.135	0.144	0.162
2@1111	Poor	0.311	0.284	0.275	0.273	0.295

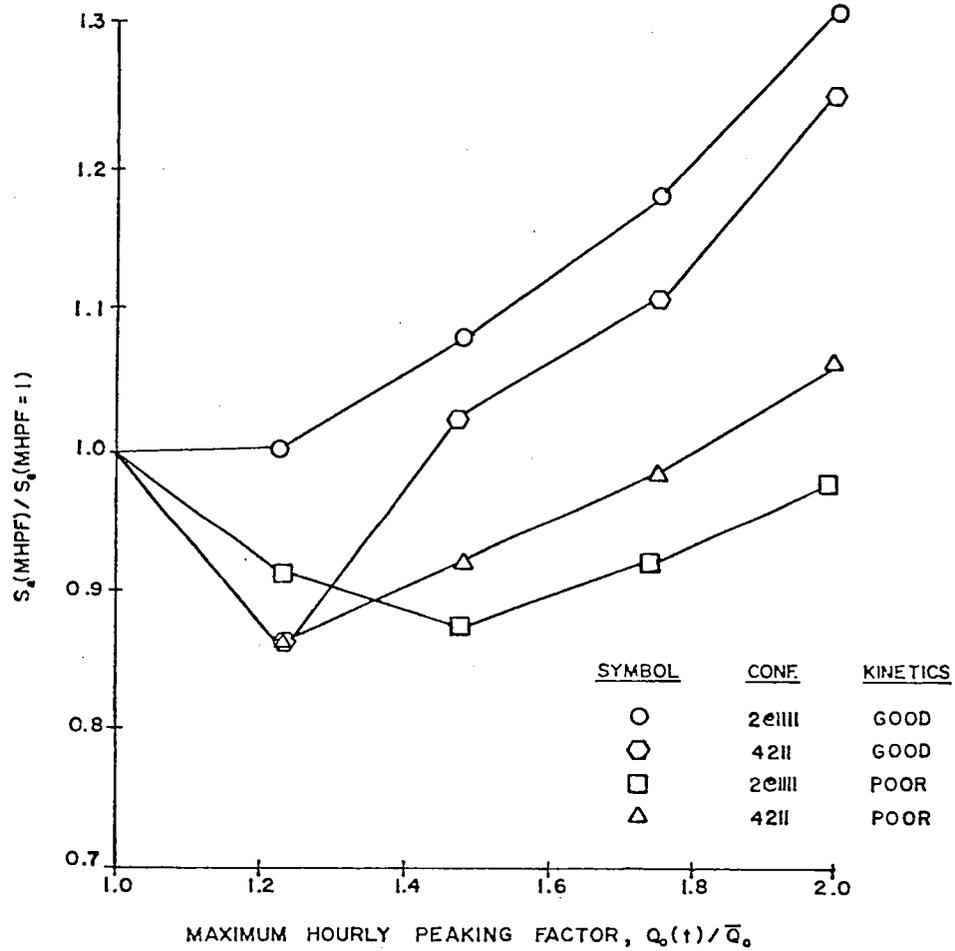


Figure 11. Effect of Peaking on RBC Performance

systems in terms of the stage psuedo-half velocity coefficient or K , and the corresponding stage influent SBOD concentration. If the influent SBOD for a stage is of the same order as K for that stage, then peaking will have more effect, in general, than if the influent BOD is an order lower than K .

The Effect of Diurnal Flow Variation on Trickling Filter Soluble BOD Concentration

Although the physical and kinetic environment in a trickling filter is every bit as complex as in an RBC, an analytical approach that considerably simplifies the characterization of trickling filter performance has been proposed by Harremoes (5). By considering the biomass to be a homogeneous porous catalytic reactor, Harremoes has shown that, for certain assumed reaction orders in the biofilm, other related reaction orders will appear to exist overall. If the reaction within the biomass, whether substrate or oxygen-limited, is assumed to follow a Monod type expression, then the external or apparent reaction of a trickling filter can only be some order between zero and first; and, in fact, may be well characterized as either zero, $1/2$, or first order. Under these conditions, Harremoes has shown that the degree of purification (the fractional removal of substrate) in a trickling filter under substrate-limiting conditions is a function of four dimensionless groups. Three of these dimensionless groups are ratios between reaction rates of the different order reactions. The fourth is the ratio of the maximum reaction rate at zero order times the detention time in the filter divided by the inlet substrate concentration. In effect, this dimensionless group compares the maximum amount of reaction that can occur in the filter to the inlet concentration. For a fixed inlet substrate concentration, this dimensionless group is directly proportional to the hydraulic residence time in the filter. Harremoes has shown that the transition between reaction orders can be defined on the $R_s X$ plane (where R_s is the fractional removal of substrate and X is the dimensionless residence time group just described), as functions of two dimensionless parameters Beta and Epsilon. Figure 12 shows a plot of the degree of purification vs. residence time in a filter (the $R_s X$ plane) for Epsilon equal to eight and

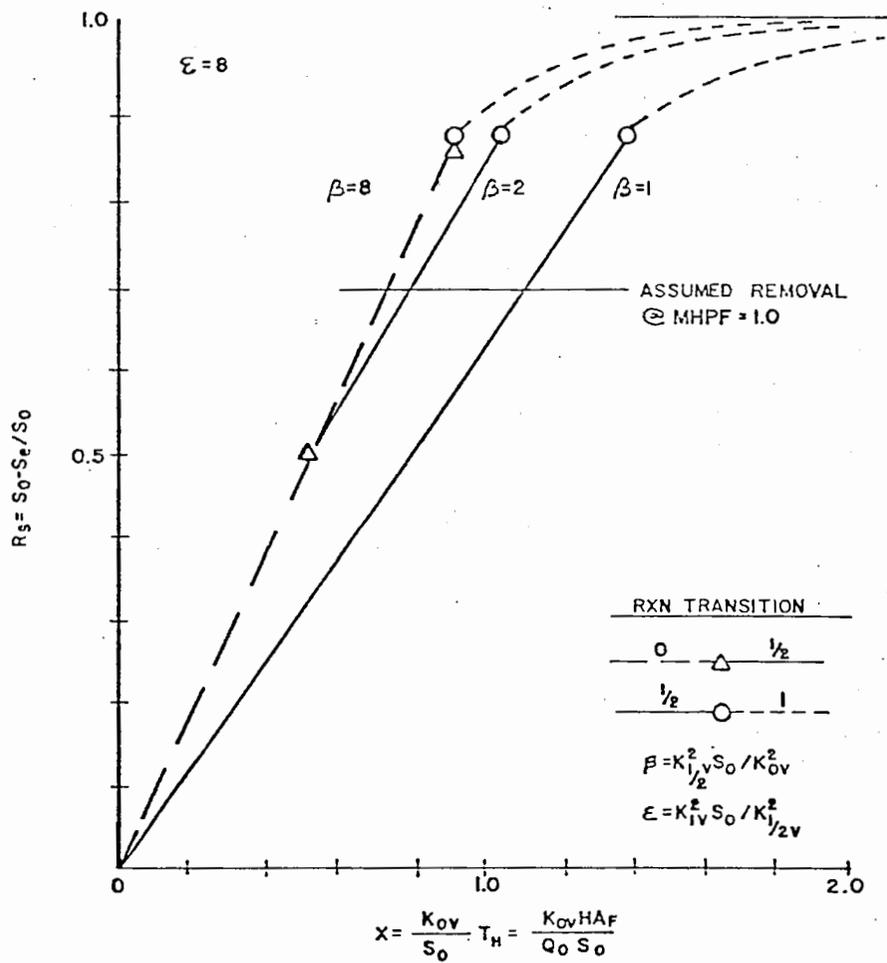


Figure 12. TF Performance (After Harremoes)

for values of Beta between one and eight. Note that when Beta is less than or equal to one, the zero order reaction will not occur anywhere in the filter. The reaction will be half order until some concentration or degree of purification is obtained and then will make a transition into first order. On the other hand, when Beta is equal to Epsilon, the half order reaction will not occur anywhere in the filter and the reaction will make a transition from zero order directly to the first order. For values of Beta between one and Epsilon, all three reaction orders may occur in the filter, depending on the residence time.

If the filter is assumed to respond instantaneously to changes in the dimensionless parameters described above, and if the kinetic coefficients remain stable, then the range of deterioration of effluent quality as a function of peaking may be calculated by a procedure exactly similar to that used for RBC simulation. Figure 13 shows the ratio of the fractional removal of soluble BOD for a given MHPF divided by the fractional removal of soluble BOD with a MHPF of one, as a function of the MHPF for the two kinetically extreme situations; one, in which Beta equals Epsilon, and two, in which Beta is equal to one. As described above, this represents the situation where the only two reactions are zero and half order, and half order and first order, respectively. In order to develop Figure 13, it was assumed that the removal with a MHPF of one was 70 percent. Here it can be seen that, depending upon the reaction order and the kinetic coefficients, the deterioration of performance by peaking is significant. These data were also plotted as the ratio of soluble effluent BOD for a given MHPF divided by the soluble effluent BOD with a MHPF of one. Figure 14 shows these data with the data from Figure 11. It is interesting to note that the two cases selected for the trickling filter closely track the results for the range of configurations selected for the RBC system. That is to say, when the reaction order is either one half or first, the deterioration of trickling filter performance is well described by the plot for the 4211 RBC configuration with good kinetics. If only zero or first order reaction occurs, then the trickling filter performance is well described by the plot developed for the RBC system with the 2 @ 1111 configuration. This would appear sensible as the trickling filter with only the half and first order reactions could be considered to be, in terms of kinetic

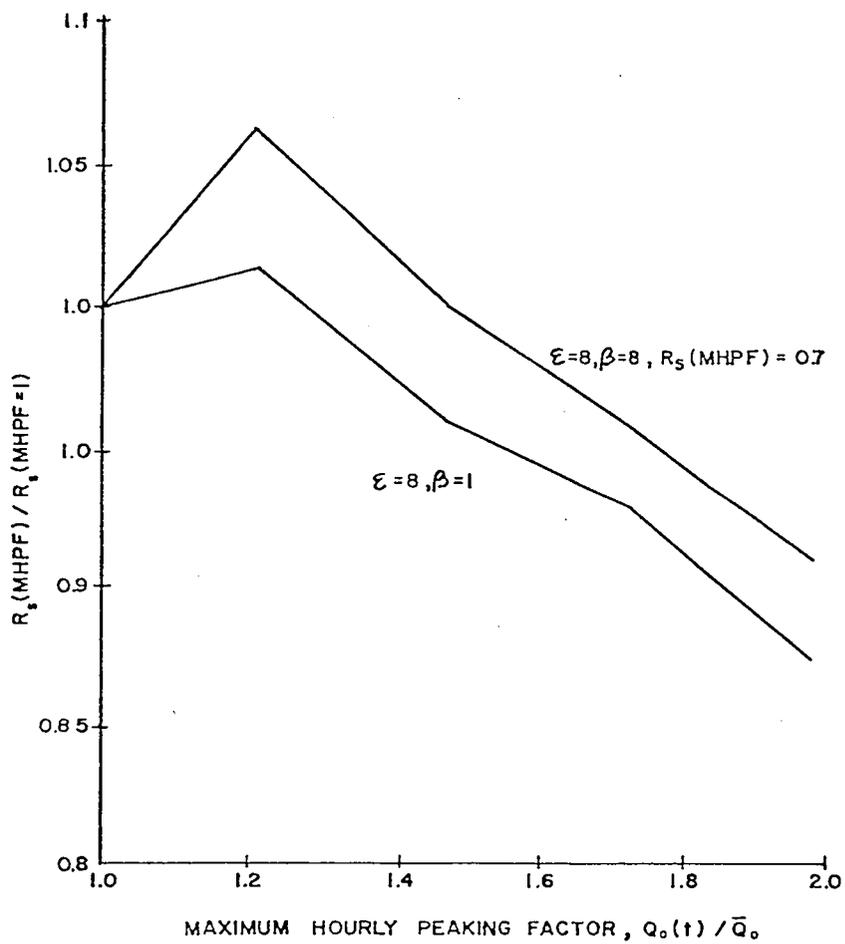


Figure 13. The Effect of Peaking on TF Performance

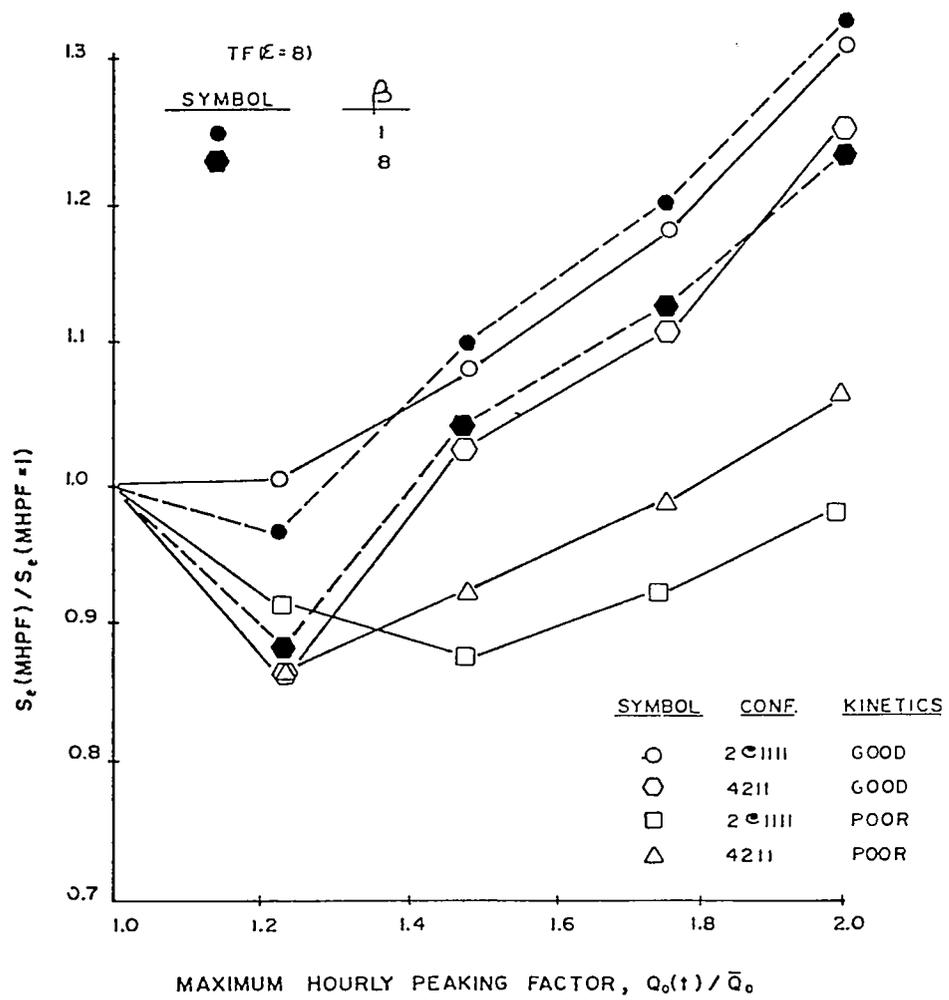


Figure 14. Comparison of the Effect of Peaking on RBC and TF Systems

potential, a more lightly loaded system, which is perhaps analogous through the more lightly loaded first stage of the 4211 configuration when compared to the 2 @ 1111 configuration.

In conclusion, the analysis developed by Harremoës would appear to be extremely useful for determining the effect of hydraulic flow variation on a trickling filter. The difficulty, of course, is in evaluating the dimensionless parameters X, Beta, and Epsilon in order to determine this degree of purification. Also, the plot assumes substrate limiting conditions either by transport or by reaction. In the event that oxygen transport or reaction is limiting, the actual performance of the trickling filter would certainly deviate from that predicted by Figure 12.

Reactor Sizing Strategy

Although it is possible to estimate the deterioration of effluent quality or at least the range of deterioration of effluent quality for RBC and TF systems due to hydraulic variation, it is significantly more hazardous to estimate the increased size of the reactor to avoid such deterioration. The reason for that is the range of reactor size increase needed to avoid deterioration may be somewhere between 0 and 50 percent, for "normal" flow variation. Also, the conditions under which it would be prudent to have additional reactor media available are not easily distinguished from the conditions in which it is not necessary to have the additional media. In the absence of methods to characterize the reaction regime for both RBC and trickling filters, it is not recommended at this time that the designer add media in order to compensate for peaking flows unless circumstances are such that the reaction regime of the attached growth process is well described.

RECIRCULATION RATIOS AND CONTROL STRATEGY

Figures 7 and 12 show the range of SBOD removal that could occur as a function of hydraulic loading and influent substrate concentration for RBC and TF, respectively. The effect of recirculation is to alter the influent

concentration, but also to increase the hydraulic loading. If the kinetic properties remain constant during recirculation, the calculations used to develop Figure 7 clearly indicate that the deterioration of performance caused by the increase in stage hydraulic loading will not be compensated for by the increase in performance caused by dilution of the influent. A critical assumption, however, is that the kinetic properties remain constant.

As the influent BOD concentration increases, the likelihood that oxygen-limiting conditions (either by reaction or transport) will encourage the growth of different types of bacteria, with different kinetic properties, is quite high. Therefore, although Figure 7 does not indicate that recirculation is a beneficial operation for RBC's, the possibility of a shift in biological speciation must be carefully considered if that shift is caused by oxygen-limiting conditions. The species that tend to be favored under oxygen-limited conditions will probably exhibit reduced ability to metabolize the organics found in domestic sewage. Therefore, performance will deteriorate due to a deterioration in the kinetic condition.

For a trickling filter system, the performance or fractional removal is a function of the dimensionless parameters Beta, Epsilon and X which are themselves functions of S_o . To illustrate the theoretical effect of recirculation, consider a condition, shown in Figure 15, in which S_o is halved by recirculation. Assume also that originally Beta was equal to 2, Epsilon was equal to 8, and X was equal to 0.7. If S_o is halved by doubling the flow, then X does not change as it is a function of the product $Q S_o$ (the kinetic coefficients are assumed not to change). As S_o is halved, Beta will change from 2 to 1 and Epsilon from 8 to 4. The 0 to 1/2 reaction order to first order reaction transition points are a function of both Beta and Epsilon, and will change as shown on Figure 15. The overall effect of recirculation will be to move from point A on Figure 15 to point B, which for the assumed conditions is a reduction in the effluent quality. Before, the fractional removal was 0.65 and recirculation has reduced this to 0.53. Therefore, while reduction in the effluent concentration alone would cause an improvement in the performance of the trickling filter, the increase in hydraulic flow counter balances that improvement (as long as the kinetic coefficients remain constant).

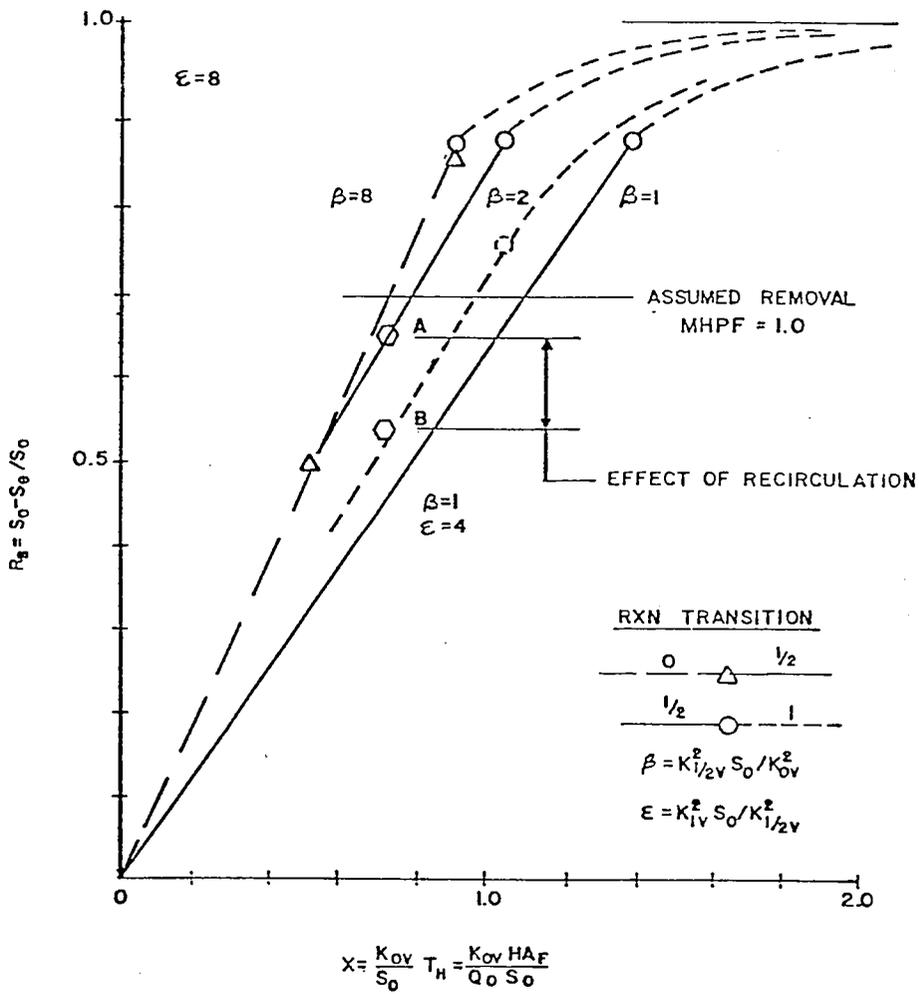


Figure 15. The Effect of Recirculation on TF Performance (Substrate limited conditions)

It must be remembered, however, that Harremoes' analysis is for substrate-limiting conditions. If the soluble BOD level is such that oxygen is limiting the performance of the filter, then transition from an oxygen-limiting condition to a substrate-limiting condition may, in effect, represent a substantial improvement in performance due to recirculation. This is included in the caveat above where it was presumed that the kinetic coefficients did not change.

Recirculation strategies, therefore, must be founded upon the actual kinetic regime within the attached growth process. If the kinetic regime is substrate-limited, then there will be little benefit from recirculation to decrease the influent concentration. For RBC and TF systems, the performance of the attached growth process is more sensitive to hydraulic loading than to influent substrate level. In fact, the higher substrate level causes a faster diffusion or transport of substrate into the biofilm and potentially higher reaction rates. In oxygen-limited conditions, however, the rate of reaction is limited by either the arrival of or the concentration of dissolved oxygen at the reaction site. If recirculation moves the device from an oxygen-limited state to a substrate-limited state, then it is possible that considerable benefit may accrue. Unfortunately, existing theory, while in a stage of rapid development, is not adequate to predict the point of transition from oxygen to substrate limitation for large scale systems.

OPERATIONAL FACTORS THAT COULD LIMIT PERFORMANCE RELATED TO PEAK FLOWS

Based on the assumptions herein, it appears that in most cases, effluent quality will deteriorate as the hydraulic loading increases, and that the change in performance due to changes in the influent substrate concentration for the same hydraulic loading depends upon a number of physical and kinetic factors. It also appears that an abrupt transition in performance may be realized if the reaction regime moves from well within the substrate-limited region to well within the oxygen-limited region. At the transition point, obviously, the effluent quality is unaffected, but as the reactor moves into the oxygen-limited regions, the performance may deteriorate

significantly compared to that expected if it remained within a substrate-limited region. The addition of returned streams such as those from sludge treatment or dewatering, or abrupt changes in the hydraulic conditions such as those caused by periodic sludge wasting, can only diminish the performance of the reactor. Under the best conditions, the deterioration in performance may be slight and proportional to the increase in hydraulic loading. However, depending upon the physical and kinetic conditions which describe performance before the addition of a side stream or return stream, the deterioration may be very significant and may represent the difference between compliance and non-compliance with effluent permit conditions. A reasonable analogy is to consider the attached growth processes as plug flow activated sludge systems, one in which there is no ability to increase the oxygen supply within the reactor. If experience or calculation would indicate the inadvisability of returning streams or wasting sludge intermittently from such a system, that should be considered as sufficient reason to avoid those conditions in an attached growth process.

One area in which an attached growth process may be significantly less sensitive than a suspended growth process is the effect of temperature. Given that an attached growth process performance is controlled by the physical transport of either substrate or oxygen, then the process performance will be relatively temperature insensitive as diffusion is only a very slight function of temperature. If, however, the reaction rate is controlled by substrate or oxygen kinetics, then the performance of the unit will decrease with temperature until physical transport controls, at which point the effect of temperature will again become insignificant.

APPLICATION OF DESIGN INFORMATION

As stated before, a critical difference between attached growth reactors and suspended growth reactors is that suspended growth reactors allow independent control of the physical transport of reactants and maintenance of dissolved oxygen levels. This greatly reduces the complexity of modelling those systems. Another significant difference is that the sludge age or growth rate may be independently determined in the suspended growth systems,

which allows relatively reliable computation for prediction of effluent quality for a particular suspended growth system.

Attached growth systems, on the other hand, can be more accurately characterized only by complex and poorly described interrelationships between transport kinetic activity and growth rate of the microorganisms. Furthermore, the operator has relatively little direct control of any of the parameters, either independently or concurrently. In the absence of internal control, the operator is limited to providing some external mode of control, such as altering the angular velocity of an RBC unit; changing the hydraulic loading of either an RBC or a trickling filter unit; providing forced draft ventilation of a trickling filter or in-channel aeration for an RBC unit; or recycling effluent to decrease the influent soluble BOD concentration, if not the mass loading of organics. Based on the discussion herein, it appears that the designer should take great care in recommending a particular sequence of external controls as the phenomena which control the internal activity of the attached growth system are not well described.

One of the greatest concerns in an attached growth system is that variation in hydraulic or organic loading may cause a variation in microbial speciation attached to the medium. While the discussion herein may provide some guidance on the response of attached growth systems to diurnal flow variation, it has been presumed throughout that the microbial speciation is constant. Based on observations of RBC installations in particular, it is clear that under low DO conditions, in the presence of sulfides, bacteria with undesirable organic oxidation properties may dominate one or more stages of an RBC system to the detriment of system performance. The significance of this caveat cannot be overstated. Therefore, while a meaningful theoretical analysis of the attached growth systems is available (at least for the steady state) and has been described herein, the designer must be constantly aware of the possibility that a significant shift in the microbial population of the system may occur with a corresponding significant change in effluent quality. Such a change in performance and effluent quality cannot be predicted by the analysis presented herein unless the alteration of the kinetic properties or the relative alterations in kinetic properties can be defined.

ACKNOWLEDGEMENT

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APPENDIX A

NOTATION

A	= Attached growth process media surface area, L^2
A_f	= Cross-sectional area of empty trickling filter, L^2
D_c	= Diffusion coefficient of substrate, L^2T^{-1}
D_w	= Diffusion coefficient of water, L^2T^{-1}
H	= TF depth of media, L
K	= Psuedo half velocity coefficient for RBC analysis, ML^{-3}
K_s	= Half velocity coefficient, ML^{-3}
K_{ov}	= Volumetric zero order reaction rate coefficient, $ML^{-3}T^{-1}$
$K_{1/2v}$	= Volumetric half order reaction rate coefficient, $M^{1/2}L^{-1/2}T^{-1}$
K_{1v}	= Volumetric first order reaction rate coefficient, T^{-1}
L_1	= Biofilm surface stagnant liquid layer depth component, L
L_2	= Biofilm surface stagnant liquid layer depth component, L
P	= Psuedo maximum rate of substrate utilization for RBC analysis, $ML^{-2}T^{-1}$
Q	= Flow rate, L^3T^{-1}

Q_o	= Influent flow rate, L^3T^{-1}
\bar{Q}_o	= 24 hr. average influent flow rate, L^3T^{-1}
$Q_o(t)$	= Instantaneous influent flow rate, L^3T^{-1}
R	= Substrate utilization rate, $ML^{-2}T^{-1}$
R_s	= Degree of purification, dimensionless
S	= Substrate concentration, ML^{-3}
S_e	= Effluent substrate concentration, ML^{-3}
S_i	= Effluent substrate concentration at the ith RBC stage, ML^{-3}
S_o	= Influent substrate concentration, ML^{-3}
T_h	= Empty filter residence time, T
X	= TF reaction time parameter, dimensionless
X_c	= Volumetric fixed film biomass concentration, ML^{-3}
f	= Fractional SBOD remaining, dimensionless
k	= Maximum rate of substrate utilization, T^{-1}
t	= time coordinate, T
α	= one half to zero reaction order ratio, dimensionless
β	= first to one half reaction order ratio, dimensionless

IMPORTANCE OF ECOLOGICAL CONSIDERATIONS ON DESIGN
AND OPERATION OF TRICKLING FILTERS

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INTRODUCTION

The effectiveness of self-purification mechanisms in the natural aquatic systems has become of major concern in Germany mainly because of the increasing need for reclaimed surface water. Thus, water pollution control regulations have been strengthened during the last couple of years. The limits for the effluent Biochemical Oxygen Demand (BOD₅) to be met by the wastewater treatment plants have been decreased from 30 mg/l to 25 mg/l to currently 20 mg/l; a further decrease to 15 mg/l is being discussed. In addition, regulations have been enforced to control the effluent Chemical Oxygen Demand (COD), suspended solids, and heavy metals. And, under particular circumstances nitrification, even denitrification and phosphorus elimination is required.

In the attempt to meet those standards wastewater treatment plants were upgraded but usually only in size, following the traditional concept of "the larger, the better". However, with changes in the overall economic conditions, the operation and maintenance costs have become increasingly important. This has

provided a strong impetus for the development of alternative concepts, or in some instances, a renaissance of almost forgotten concepts.

Currently, two-stage biological systems are enjoying increasing attention. Especially, the combination of an activated sludge treatment process as a first stage, and a trickling filter for secondary biological treatment has been discussed in reference to a number of older treatment plants in Niedersachsen, Germany (1) as well as to the treatment plant at Lima, Ohio (2). This particular concept has been supported recently by the excellent and highly reliable treatment achieved at the full-scale plant operated by the City of Lahr, Germany (3).

To be able to exploit the potential inherent in this particular process combination it is essential to address the duties to be accomplished by the consecutive system elements carefully, and in accordance with the ecological principles of microbial systems. The purpose of this paper is to demonstrate the significance of those ecologically based principles on the design and optimization of trickling filters operated as a second biological stage.

ECOLOGICAL BACKGROUND

It should be stated that the biological systems employed in wastewater treatment plants are self-adjusting in character. Thus, the organism-type distribution in any reactor is the result of the particular combination of environmental factors in effect. Examples of these factors are the presence of particular nutrient substrates, substrate concentration, oxygen availability, and temperature. In turn, the organism-type distribution determines the capacity of the biocommunity to metabolize particular types of pollutants such as soluble organics, organic particles or ammonia. It also determines the rate at which those pollutants are metabolized.

Any change in the combination of these factors will change the organism type distribution (Fig.1), and consequently the capacity of the biocommunity to metabolize specific substrates (Fig.2). We can conclude that optimization of biological wastewater treatment processes must be focused on the optimization of the environmental factors effective prior to any technological improvement.

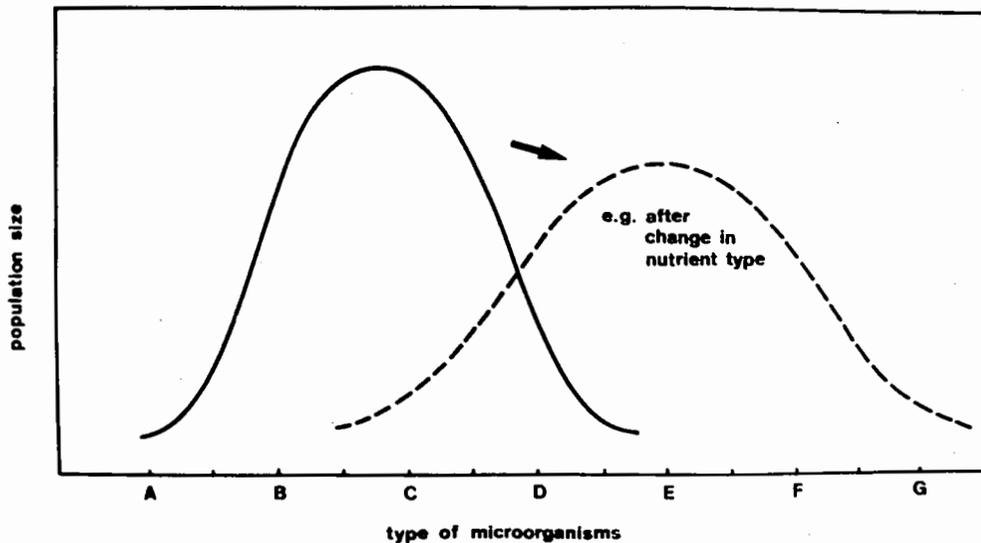


Fig.1 Effect of a particular environmental factor on the distribution of various types of microorganisms

The various organism types active in wastewater treatment systems are linked to each other by either substrate-product relationships, or predator-prey relationships (Fig.3). As a result, the composition as well as the size of the microbial population shifts as primary substrates are converted into secondary substrates. This may happen in a batch reactor, in a BOD bottle for instance, as the reaction time elapses, or in a suspended growth reactor, if sludge age is increased gradually. It happens downstream a river, and downstream in a plug-flow fixed bed reactor (4). Consecutive ecological zones appear, each zone characterized by a particular organism-type distribution as well as by a specific metabolic capacity.

The development of such sequential zones of microbial populations can easily be simulated by a reactor cascade experiment (5). Typical results of such an experiment are provided in Fig.4. Evidently, heterotrophic bacteria predominate within the very first zone of the cascade, and convert soluble organics into biomass and ammonia. Only after the concentration of soluble organics approaches the lowest level do protozoa become predominant, and simultaneously organic particles are removed. And only after this particular process is almost completed does nitrification take place.

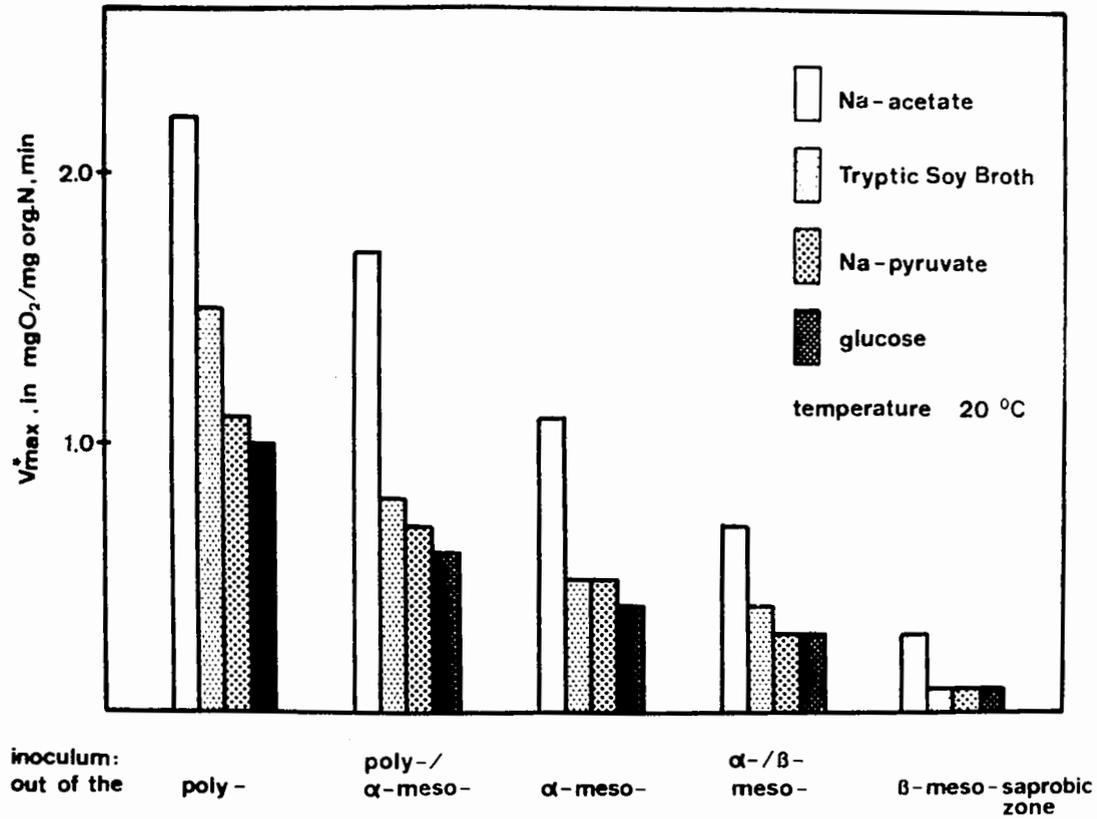


Fig.2 Metabolic capacity of microbial biocommunities (Aufwuchs) from various eutrophic zones of an initially high-polluted river

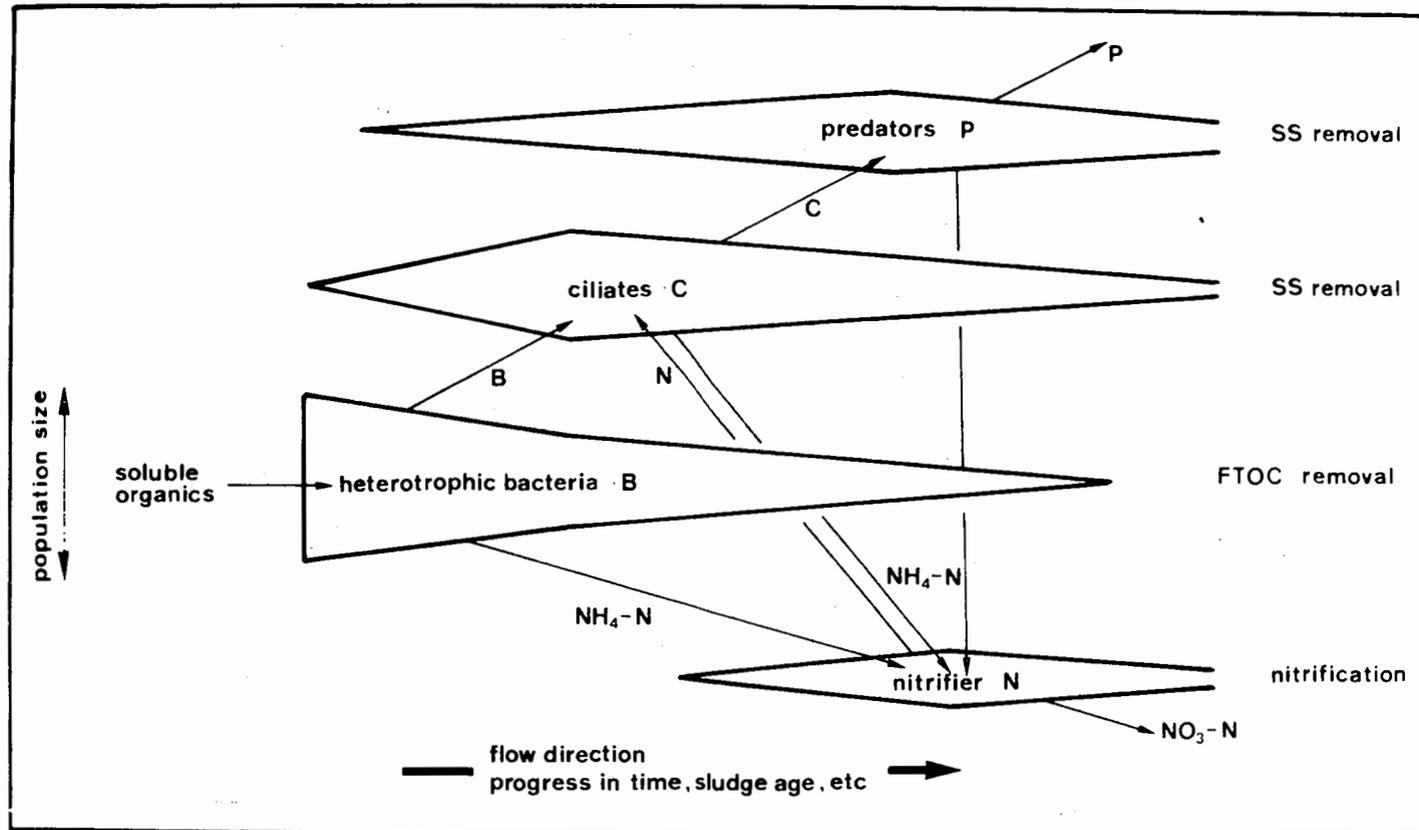


Fig.3 Principles of the succession of microbial biocommunities in aquatic systems

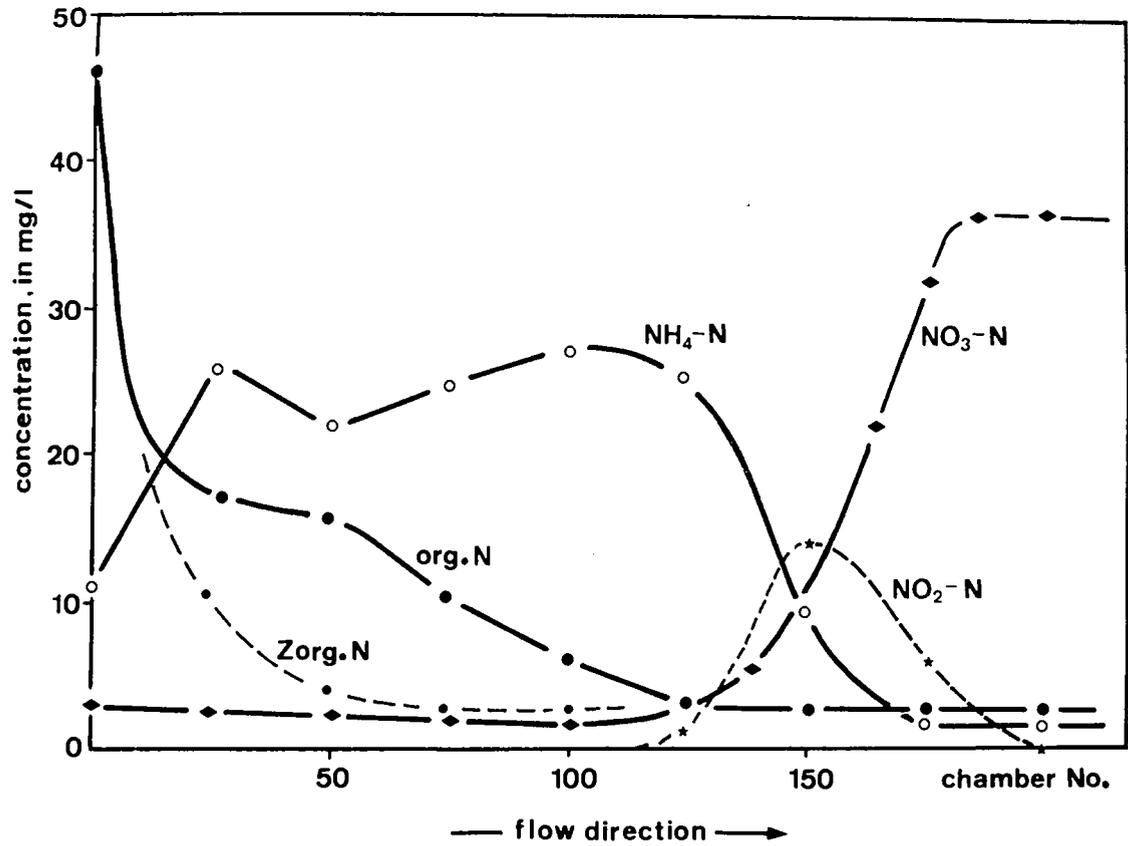


Fig.4 Effect of population shifts downstream a model river on the ammonification/nitrification capacity of the bio-communities

BIOTECHNOLOGICAL CONCLUSIONS

Since that rate at which a particular substrate is metabolized is directly proportional to the number of organisms capable of accomplishing that particular reaction (Fig.2) the establishment of conditions favoring the growth of the desired population must be judged beneficial. And because such conditions differ for each population, the establishment of consecutive sequential zones within a wastewater treatment system is essential. To achieve a high overall process economy the development of such a sequence of biocommunities should be the goal of process design.

Plug-flow fixed bed reactors in general provide the prerequisites for the development of microbial succession. Attached to the filter medium the organisms are fixed in position. The water flowing through carries substrates and, in turn, products to the consecutive biocommunities.

In general, a complete succession beginning with a heterotrophic bacterial population can be established in such reactor types. The degree of succession achieved depends more or less only on the filter depth provided, and on the possibilities for maintaining aerobic conditions. From a practical point of view, however, such a single stage reactor is disadvantageous because of the well known problems associated with excess growth of the heterotrophic bacteria in the upper zone of a plug-flow fixed bed reactor. Those problems can easily be diminished by splitting the process into two stages with the first reactor designed for the development of a heterotrophic bacteria community, and the consequent removal of soluble organics. The biomass produced by this process should be separated by an intermediate clarifier to decrease the suspended solid loading of the second biological stage. The duty of this secondary reactor is to remove the remaining suspended solids by protozoa activity, and to remove the ammonia by nitrification.

To operate such a two-stage system economically it is important to know about the pre-treatment effect which must be provided by the first stage process in order to get the second stage system to operate as desired. And, under any chosen pre-treatment efficiency, the effects of operation parameters such as hydraulic loading, type of filter media, filter depth, temperature, and so forth must be investigated. Such studies are currently being conducted at the Karlsruhe wastewater treatment plant. First results can be provided herewith.

PILOT PLANT

A sketch of the pilot plant used for the investigations is presented in Fig.5. Four systems were operated in parallel to study the effects of parameter variation under identical conditions (identical raw wastewater characteristics and temperature for four different loadings). Details characterizing the system elements of the pilot plant are listed in Tab.1.

Tab.1 Design and operation data for the pilot plant

primary clarifier	volume	14.8 m ³
	inflow rate	17 m ³ /h
aeration tanks	volume	2.5 to 4.5 m ³ (adjustable)
	inflow rate	0.85 to 3 m ³ /h
clarifiers	volume	4.5 m ³
rock media trickling filters	diameter	1.2 m
	depth	2.5 m (A) 3.5 m (B)
	medium	lava rocks 16/40 mm
	inflow rate	0.4 to 1.1 m ³ /h
plastic media filter	diameter	1.2 m
	depth	2.5 m
	medium	Flocor R, ICI (corrugated hose pieces, 35mm)
	inflow rate	0.4 to 1.1 m ³ /h

Raw wastewater was used as feedstuff. The wastewater was pumped, in excess, to a primary clarifier. The characteristics of this primary effluent are listed in Tab.2. Portions of the clarifier effluent were then distributed to the aeration tanks. Flow rate and aeration tank volume were varied in order to adjust the hydraulic loading of the system to the desired values.

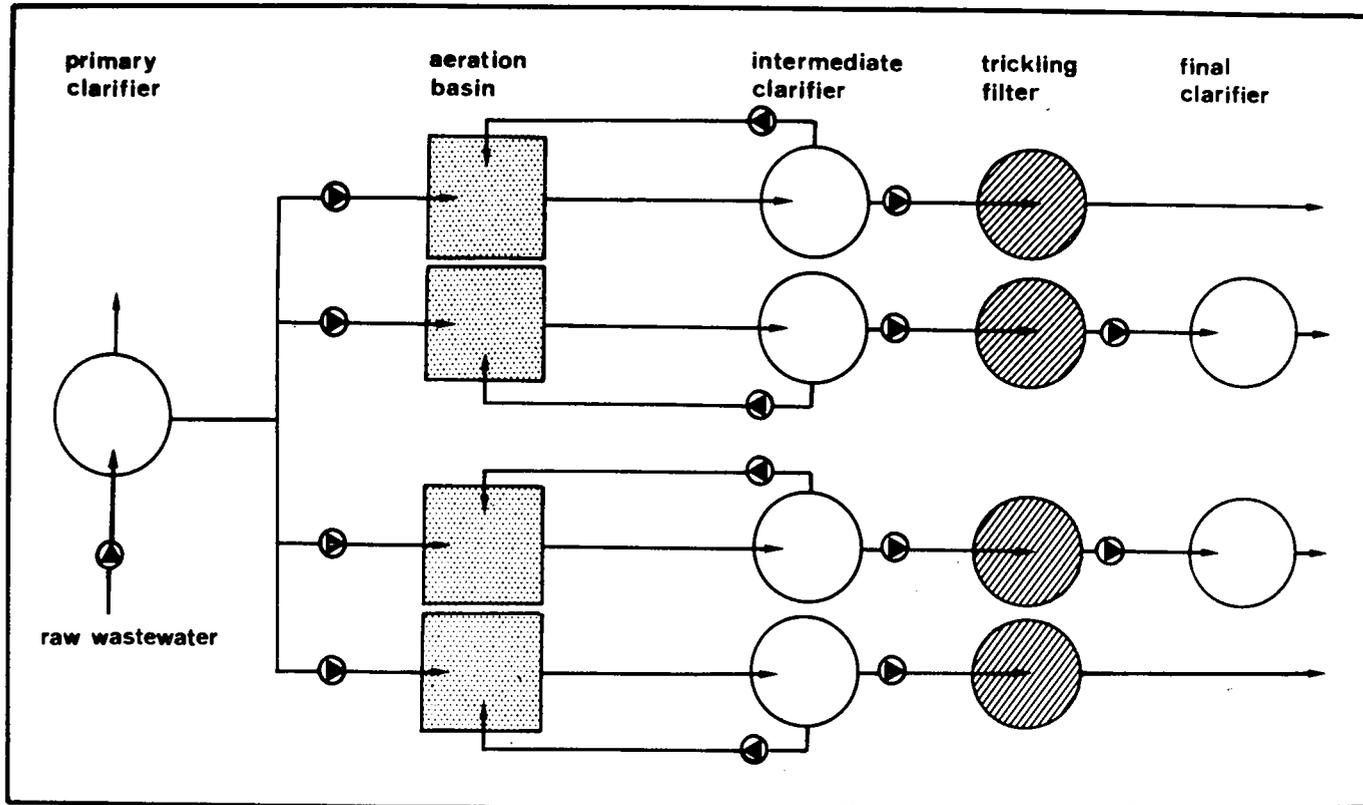


Fig.5 Schematic flow diagram of the pilot plant operated at the Karlsruhe wastewater treatment plant

Tab.2 Characteristics of the primary effluent

<u>parameter</u>	<u>range</u>	<u>average</u>
BOD ₅ ,in mg/l	100 - 220	170
COD ,in mg/l	280 - 500	380
FTOC ,in mg/l	25 - 75	45
NH ₄ -N ,in mg/l	8 - 25	20

Trickling filters were used as the second-stage biological reactor. Each filter was equipped with sampling sites every 50 cm down the columns. Thus, the partially treated wastewater as well as organisms attached to the filter media could be taken for profile measurements.

The entire pilot plant was covered by a tent in order to make the investigations less dependent on climatic conditions.

RESULTS

During the two year experimental program the substrate loading of the pretreatment system (i.e. activated sludge treatment process) was systematically changed in a range between 0.24 and 2.63 kg BOD₅/kg MLSS,d. As expected the concentration of soluble organic substances in the effluent of the intermediate clarifier increased, but only slightly, with increasing F:M ratio. In contrast to this, the concentration of the effluent suspended solids increased significantly, causing a considerable increase in the overall BOD₅ (from about 12 to 78 mg/l), as well as in the overall COD (from about 54 to almost 200 mg/l).

As mentioned above, the organic suspended solids washed out of the intermediate clarifier serve as a potential substrate for protozoa. Thus, it was expected that protozoal activity would become predominant in the upper zone of the trickling filter.

Fig.6 illustrates the typical results of the protozoal activity. Under the particular influent conditions, almost 70 percent of the influent COD were removed within the first 75 cm of the filter. Compared to this, the contribution of the organisms settling in the deeper regions of the trickling filter was almost negligible.

Comparison of filterable COD (FCOD) and total COD proved that the mechanisms responsible for the COD removal in the upper zone were characterized more by protozoal than by bacterial activity.

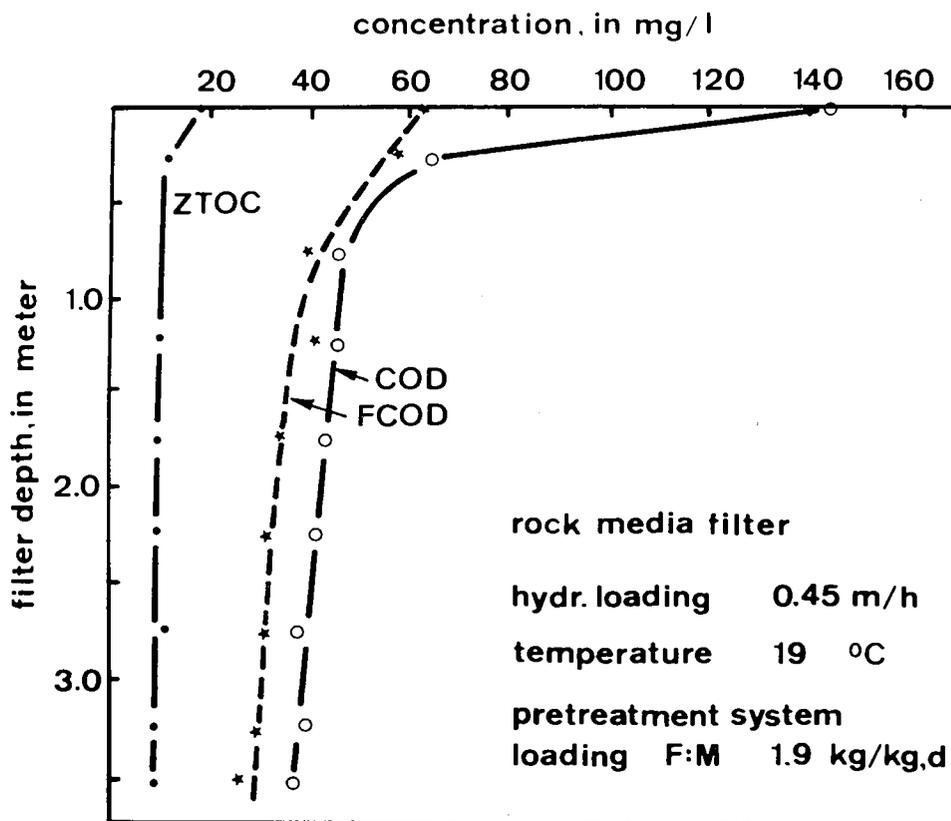


Fig.6 Typical concentration profile for the trickling filter. Parameter: organic pollutants

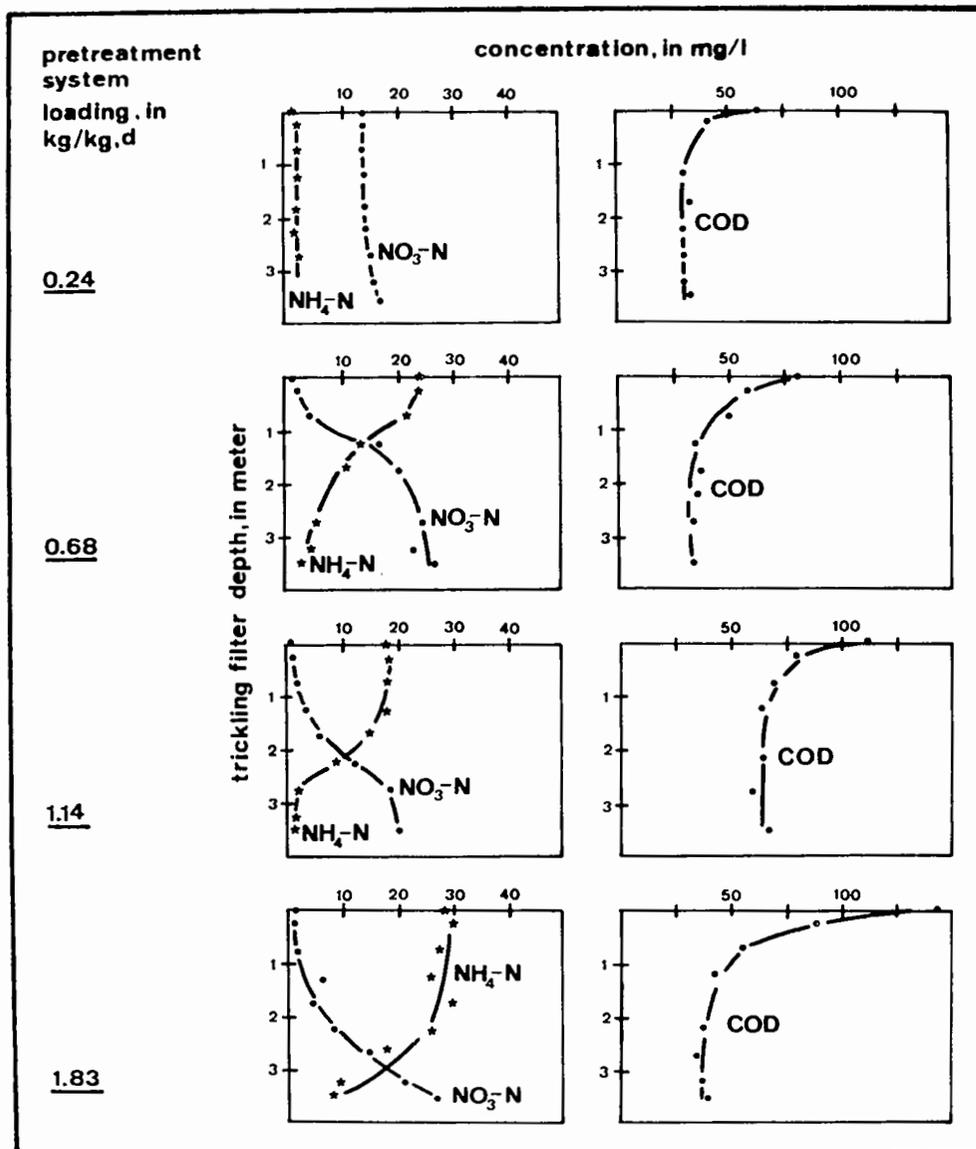


Fig.7 Effect of the efficiency of the biological pretreatment system on the trickling filter performance.
 Operation criteria: rock media filter,
 hydr.loading 0.75 m/h
 temperature 13.5 °C

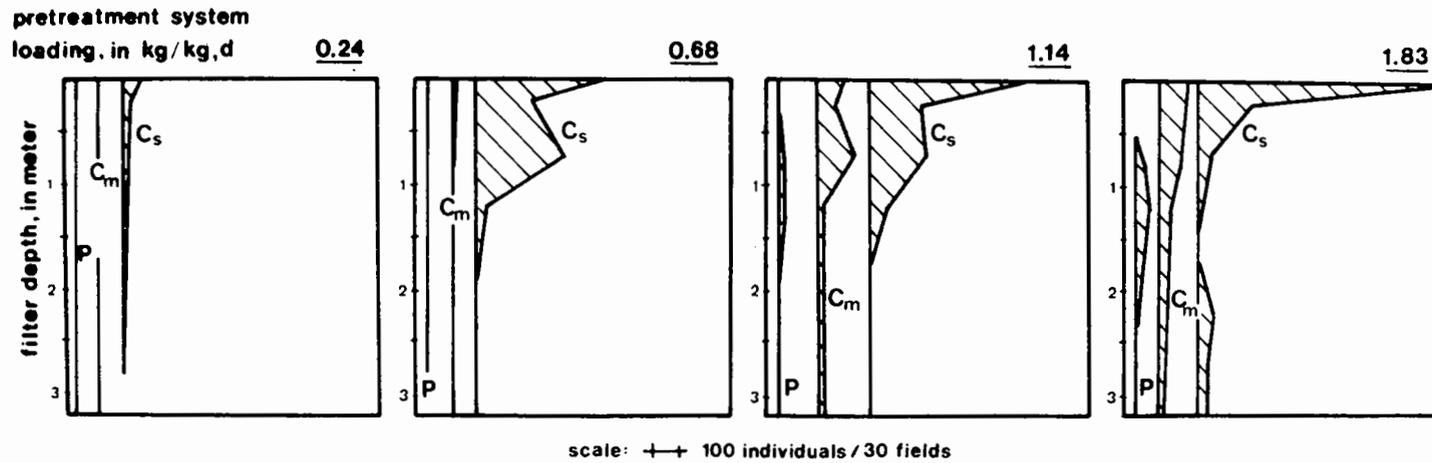


Fig.8 Effect of the efficiency of the biological pretreatment system on the distribution of sessile ciliates (C_s), motile ciliates (C_m), and predating ciliates (P).

For details see Appendix A and B

With decreasing influent COD caused by the decreasing loading rate of the pretreatment system the nutrient supply for protozoa and associated heterotrophic bacteria became poor. The zone responsible for COD removal decreased (Fig.7) as the habitats changed, and consequently the diversity and number of protozoa as well as their predators decreased (Fig.8; App.A,B). Apparently, this is correlated to the development of nitrifiers. Nitrifiers appear predominantly because there is less competition for the available oxygen once the heterotrophs have gone.

The importance of the latter statement can be experimentally demonstrated by adding an organic substrate to the trickling filter influent, in excess concentration, for a short period of time (shock loading). The effects are illustrated in Fig.9.

The presence of easily degradable substances within the nitrification zone obviously inhibits nitrification, at least in the inner region of the trickling filter. Close to the effluent region where the overall population density is diminished, and where the oxygen supply is less dependent on the trickling filter ventilation efficiency, nitrification is less inhibited. These differences can be satisfactorily explained by the competition for oxygen by the various species forming the bio-community, with nitrifiers as the most sensitive competitors. They are the first to be affected by a decrease in available oxygen.

Of course, shock loading situations affecting the second stage of a two-stage process are rather academic in nature, but oxygen deficiency effects are certainly not.

Those trickling filters which were loaded heavily with suspended matter over a long period of time (pretreatment process loading above 1.2 kg/kg,d) gradually became less efficient as far as nitrification is concerned. As demonstrated in Fig.10 the decrease in ammonia concentration slowed down significantly after a filter passage of about 2 meters. The inefficiency of the natural ventilation combined with partial clogging of the filter pores may have caused this effect. When the filter was artificially aerated, the nitrification efficiency was significantly improved. An even stronger beneficial effect on nitrification was noted when the filter was flushed with water once a week.

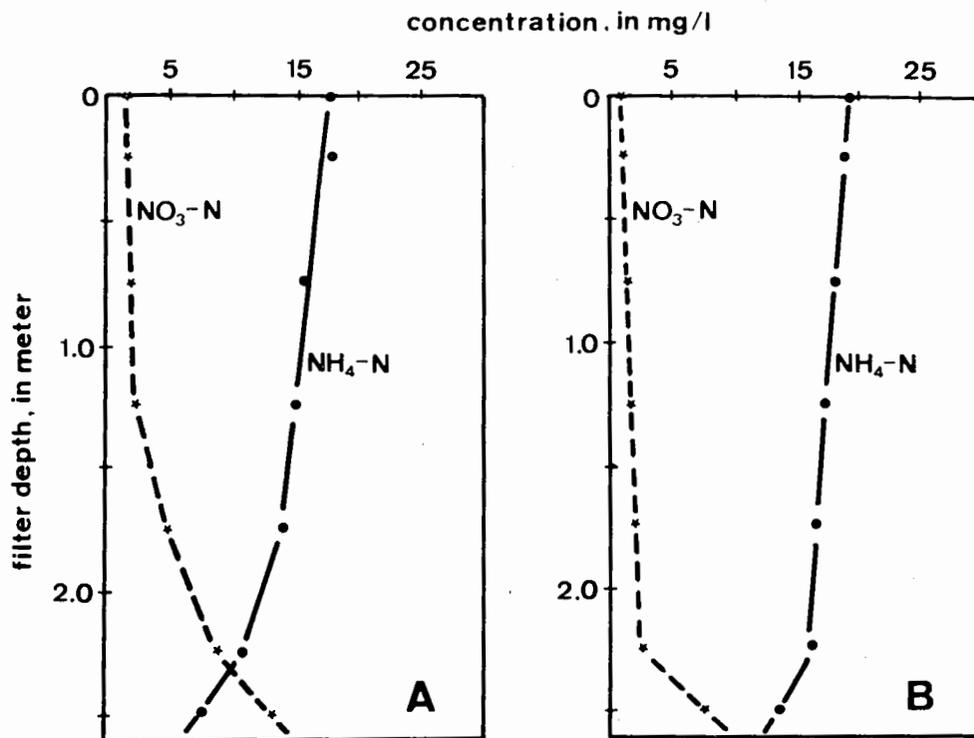


Fig.9 Effect of organic substrate shock loading on the performance of the trickling filter

- A: initial state
influent/effluent FTOC: 48/31 mg/l
- B: Na-acetate, added to the trickling filter influent
for 3 hours
influent/effluent FTOC: 230/166 mg/l

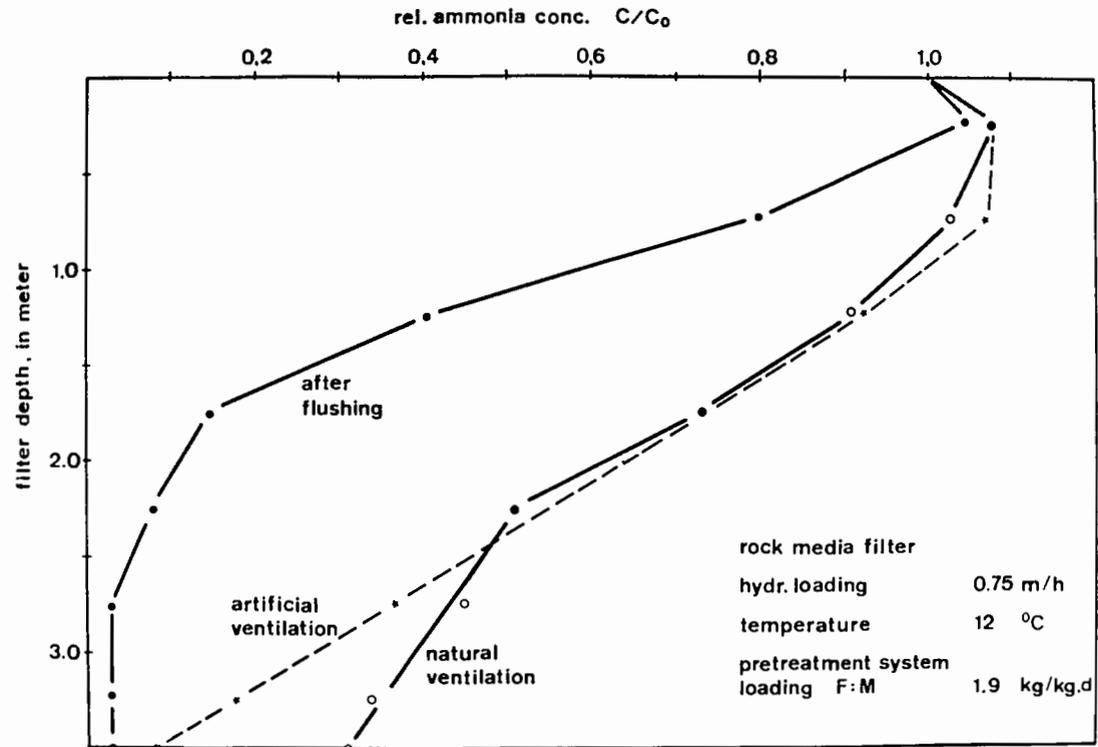


Fig.10 Effect of artificial ventilation, and occasional flushing respectively on the nitrification efficiency

After analyzing the technological consequences of these results the replacement of the filter media by one's which would permit a better air circulation was considered. Flocor R, ICI was chosen because this material offers a surface area similar to the rock type media previously used, but has a much greater pore volume.

In order to compare the efficiency of both filter media, a rock media and a plastic media trickling filter were operated in parallel. Feedstuff, filter depth, and hydraulic loading were identical. But the results gained were rather discouraging. Certainly, the COD removal capacity was almost the same, but the nitrification efficiency was not (Fig.11).

Evidently, the filter depth of the Flocor R filter must be increased significantly to achieve comparable nitrification effects. This in turn, has the negative effect of increasing the energy costs for pumping.

Anyway, the question to be answered is: Why do these differences in efficiency occur? The detention time distribution functions for both filters in comparison (Fig.12) might answer this question.

The plastic media filter detention time distribution function resembles a mixed situation rather than a plug-flow situation. Short-circuiting of the water flow is probably the major reason for this. For the development of ecological zones such hydraulic conditions are disadvantageous. Organic material is imported to the deeper regions of the trickling filter, and generate a competitive situation as described previously and illustrated in Fig.7 and 9. Technologically, those effects can be compensated for by increasing the filter depth, if there is no other possibility for handling the hydraulic problem.

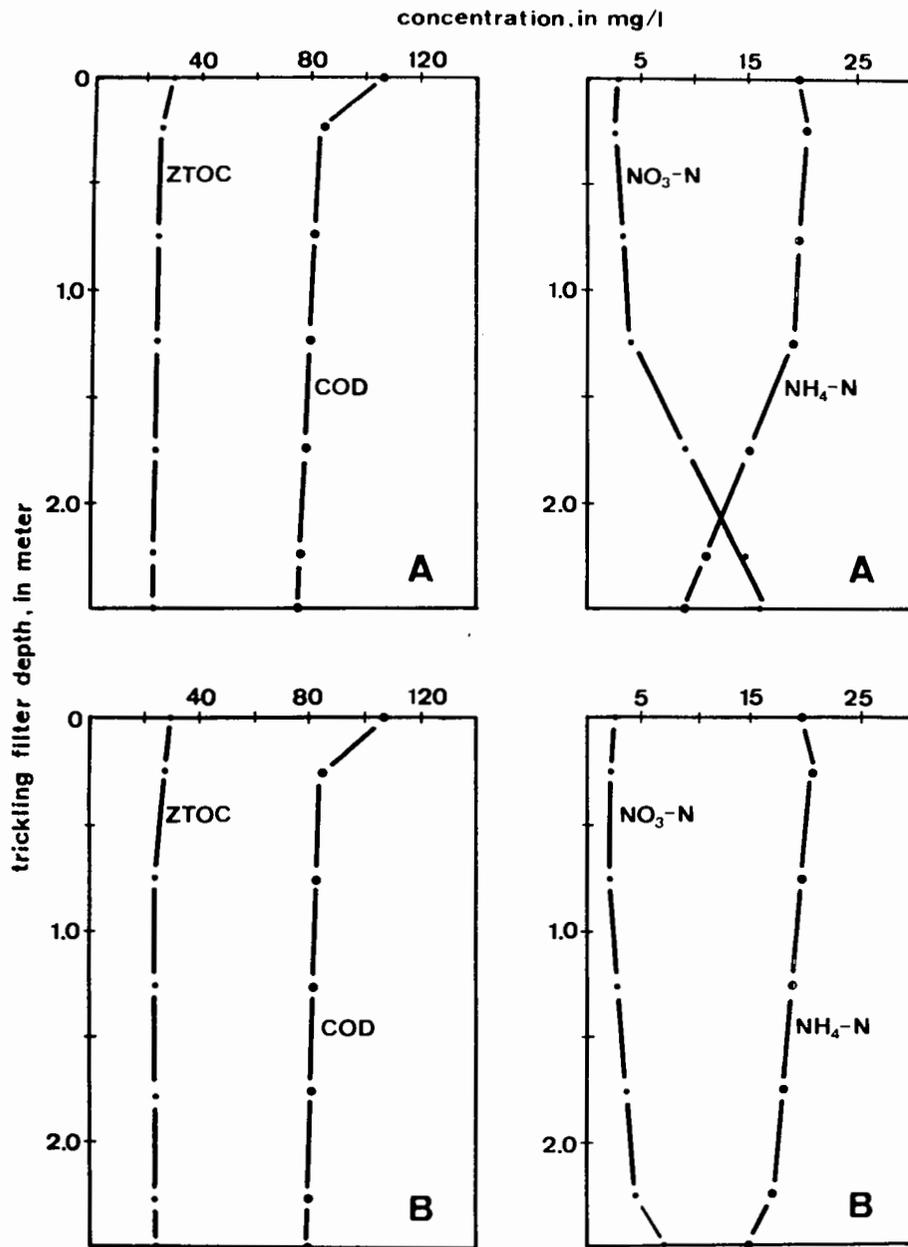


Fig.11 Rock media (A) and plastic media (B) trickling filter in comparison.
hydr.loading: 1.0 m/h

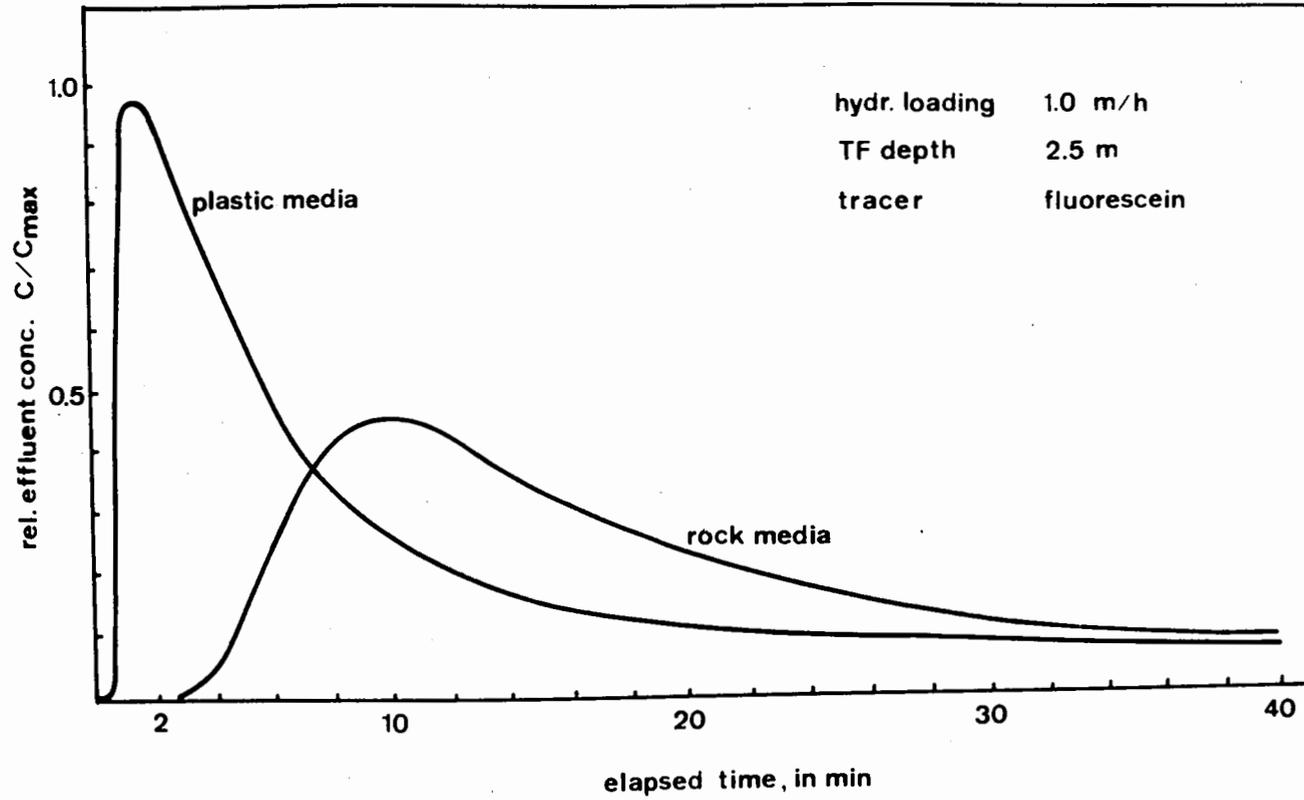


Fig.12 Detention time distribution function for the rock media and plastic media filter.
Filter depth: 2.5 m ; hydr.loading: 1.0 m/h

SUMMARY AND CONCLUSIONS

The experimental results described above support the general idea that the establishment of consecutive environmental zones within a wastewater treatment process is technologically feasible, and economically beneficial. But that is not a surprise at all since we did nothing but copy patterns typical for natural aquatic systems.

As sanitary engineers, we have always tried to copy these patterns. But with increasing emphasis on single sludge, extended aeration systems - whatever the reason for that emphasis may have been - the connection to the ecological fundamentals were lost. With single sludge systems we have attempted to solve all problems, and evidence proves we can almost do it, but we have to pay for it.

The choice of environmental factors necessarily favors only particular biocommunities, and is usually disadvantageous for the other biocommunities of concern. It is impossible to establish a biocommunity which is simultaneously as active in soluble organics removal as it is in nitrification. Thus, the system size must be enlarged, making tremendous costs, especially energy costs, inevitable.

Two-stage biological systems favoring the establishment of a succession of biocommunities for specific duties is a promising alternative.

The job of the first stage is the conversion of soluble organics into biomass, and the separation of the resultant biomass as effectively as possible. Heterotrophic bacteria are needed to accomplish this job. Thus, the system must be operated at a high loading rate to favor the establishment of those organisms.

Suspended solids and ammonia are the pollutants which characterize the effluent of such a high-loaded system. They can be removed by protozoa (suspended solids), and by nitrifiers (ammonia, nitrite). Thus, the establishment of those organisms must be the major goal of the second-stage process design. Under natural conditions these organisms grow in different but consecutive zones. Consequently, a reactor type must be provided which offers the chance for heterotrophs and nitrifiers to be established in consecutive zones.

In this context, a plug-flow trickling filter is judged to be almost ideal. Results of pilot plant studies prove that this judgement is reasonable. Plug-flow conditions and a fixed-bed for attached growth are prerequisites in the attempt to establish a succession of biocommunities within a reactor. In addition, the environmental requirements of the organisms must be met as closely as possible. The oxygen supply for the nitrifiers seems to be the most sensitive parameter. To achieve a high nitrification efficiency sufficient air circulation within the trickling filter must be maintained, and any factor causing a competition for oxygen within the nitrification zone must be minimized by process design and operation. Forward mixing, shock loading, and excess particle introduction and entrapment are major factors which must be avoided.

ACKNOWLEDGEMENT

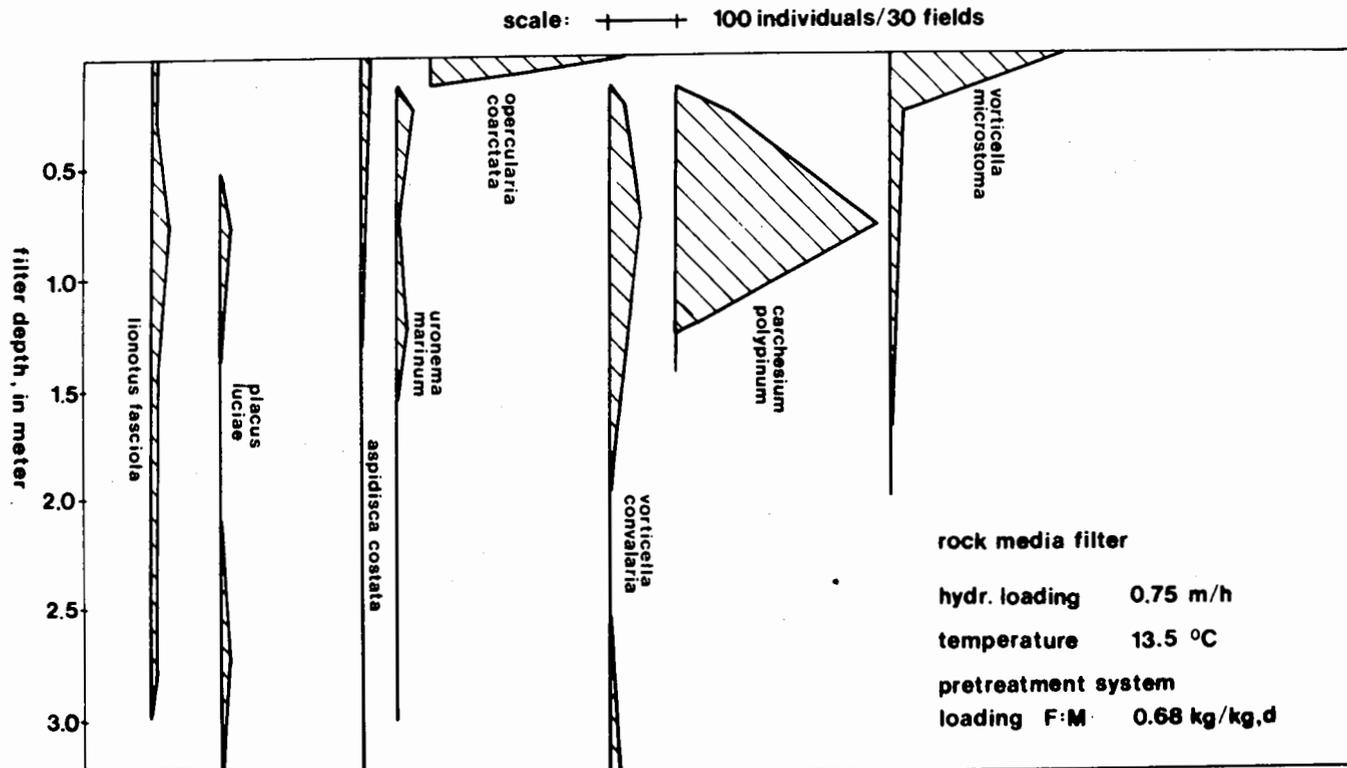
This research work was supported by Grant o2 WA 947 from the German Ministry of Research and Technology. The authors also wish to acknowledge the contribution of Günther Frietsch who did the microbial analysis. The valuable discussions with W. Segel, UC Davis are gratefully appreciated.

SYMBOLS

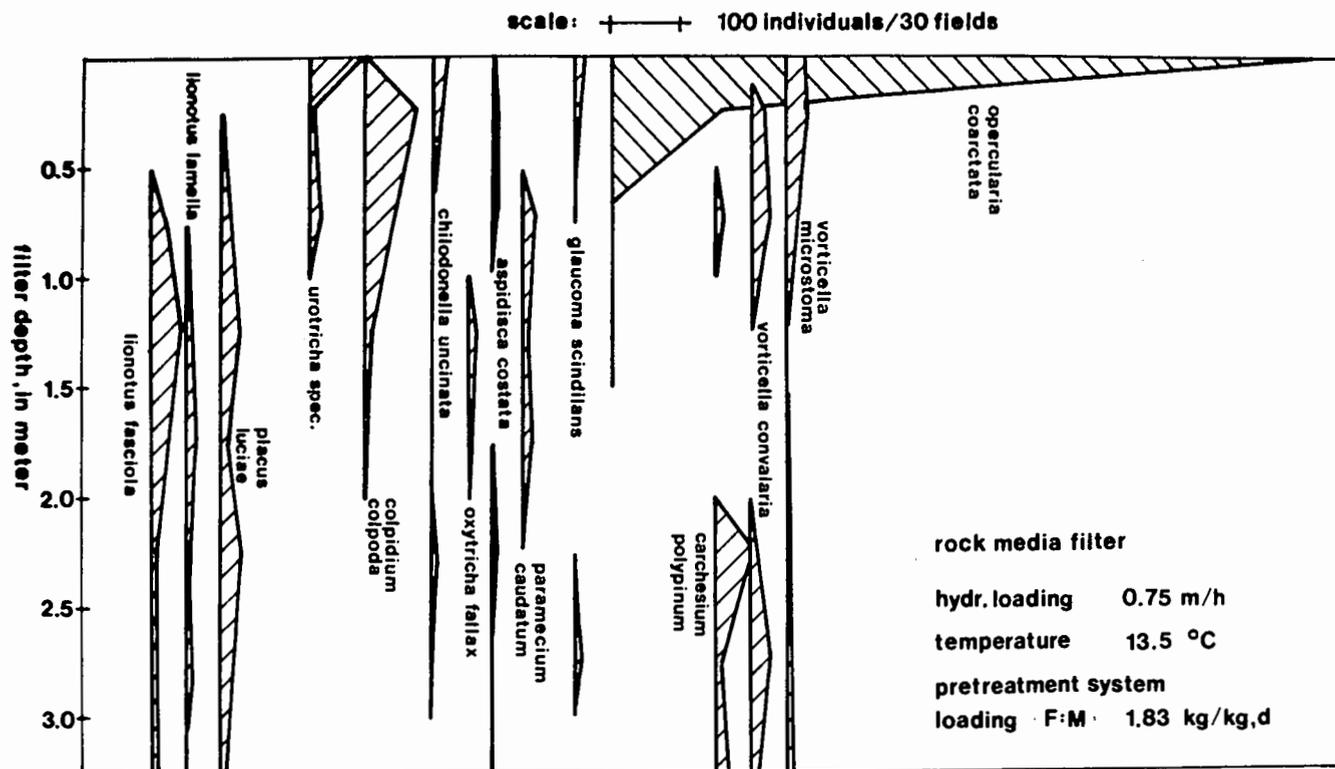
ZTOC	centrifugable Total Organic Carbon, in mg/l
F:M	Food to Microorganism ratio, in kg BOD ₅ /kg MLSS, d
FCOD	filterable Chemical Oxygen Demand, in mg/l
V _{max} *	specific maximum respiration rate, in mg O ₂ /g orgN, min

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Appendix A Distribution of ciliates within the trickling filter.
 Medium loaded activated sludge treatment system in front



Appendix B Distribution of ciliates within the trickling filter.
High loaded activated sludge treatment system in front.

EVALUATION OF BIOLOGICAL TOWER DESIGN METHODS

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INTRODUCTION

The design of biological towers (trickling filters) has been approached by the use of numerous empirical relationships. One of the first empirical relationships developed was the National Research Council (NRC) formula. It was formulated in 1946 and was based upon data collected from treatment plants serving military installations across the United States during World War II (1). It is only applicable to rock trickling filters.

Starting with the work of Velz (2) in 1948, a series of relationships were developed based upon first-order kinetic removal of BOD. The removal rate for these first-order relationships have been developed in terms of depth of filter or time of contact of wastewater with the tower media. Velz's (2) and Stack's (3) relationships are examples of the use of the first-order reaction rate with depth of tower and Howland's (4), Schulze's (5), and Eckenfelder's (6) formulations are examples of the first-order reaction rate with time of passage.

In 1970, Cook and Kincannon (7) presented data that showed the performance of a biological tower was dependent upon the organic loading (lbs BOD/day/1000 ft³) applied to the tower. This data also showed that the relationship between lbs BOD applied/day/1000 ft³ was not a first-order relationship. Between 1974-1976, Stover and Kincannon (8, 9, 10) presented data showing that the same relationship applied to rotating biological contactors. It is also interesting to note that Velz (2) also presented a very similar concept in 1948. However, his first-order relationship overshadowed his total organic loading concept. It should also be pointed out that the NRC formula is based upon a total organic loading concept.

Other models and empirical relationships have been developed, however, only the first-order kinetic relationships and the organic loading relationships will be evaluated.

FIRST-ORDER KINETIC RELATIONSHIPS

First-order kinetic relationships describing the removal of BOD by a biological tower have been accepted and used by many designers since Velz first introduced the concept. These relationships take the following forms:

$$\frac{S_e}{S_i} = e^{-K_1 D} \quad (1)$$

$$\frac{S_e}{S_i} = e^{-\frac{K_2 D}{Q^{0.666}}} \quad (2)$$

$$\frac{S_e}{S_i} = e^{-K_3 A_v^{1+m} D/Q^n} \quad (3)$$

where S_e = soluble BOD at depth D, mg/ℓ

S_i = influent BOD, mg/ℓ

K_1, K_2, K_3 = treatability factor

D = depth of biological tower, ft

Q = hydraulic loading, gpm/ft²

A_v = specific surface area of tower media/ft²/ft³

m, n = constants characteristic of media

Deviations from the accepted models are known to occur. These deviations are usually attributed to either a temperature effect or saturation of the filter. Temperature effects will be discussed later in this paper. Saturation occurs when the system becomes oxygen limiting and will be discussed in more detail in the organic loading section of this paper.

Another situation that causes deviations by the first-order models is the occurrence of more than one kinetic rate through the depth of the tower. This is shown in Figure 1. It is seen that in all but one case there are two first-order rates for each removal curve. This is also shown in Table I where the reaction rates have been determined for both phases.

Table I. Reaction Rate Constants

Wastewater	Inf. BOD ₃ mg/ℓ	Flow rate gpm/ft ²	K ₁ * ft ⁻¹	K ₂ * ft ⁻¹
Slaughterhouse	300	1.74	0.035	-
	320	1.04	0.056	0.024
	400	0.52	0.099	0.022
Pulp and paper	210	1.04	0.044	0.018
	290	0.35	0.053	0.020
	220	0.52	0.054	0.019
Domestic #1	70	1.25	0.241	0.076
Domestic #2	20	0.95	0.408	0.076
	114	1.06	0.087	0.060
Carbohydrate	30	0.52	0.274	0.072
	80	0.52	0.198	-
	110	0.69	0.197	0.085
	180	0.69	0.081	-

* Based upon natural logs

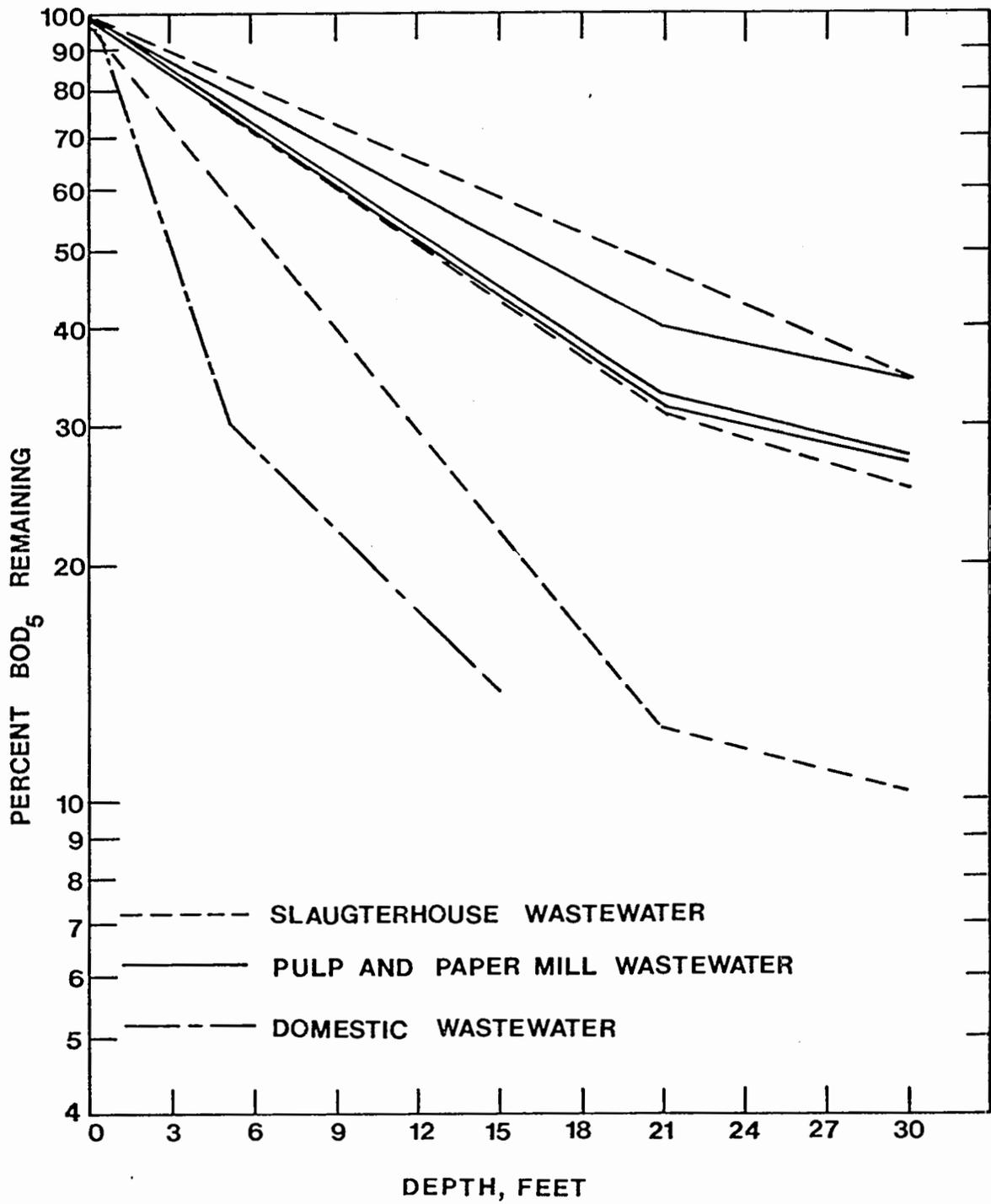


Figure 1. BOD remaining as a function of tower depth.

It is interesting to note that the phase I reaction rates vary with influent BOD, hydraulic loading, and type of wastewater. However, the phase II reaction rates vary only with type of wastewater. A phase II reaction rate of approximately 0.02 ft^{-1} was found for the slaughterhouse and pulp and paper wastewaters. A value of approximately 0.075 ft^{-1} was found for domestic and a carbohydrate wastewater.

It is seen that if the first order reaction rate with depth of tower relationship is to be used in designing biological towers, the reaction rate constant must be specific for a particular wastewater at a given influent BOD and hydraulic loading. If either of these change, then the reaction rate constant must be changed. This puts some strong handicaps on the design relationships.

The first-order reaction rate with time of passage helps to correct this problem, but does not completely solve the problem. The relationships developed to express the time of passage includes the hydraulic loading. Thus, a change in hydraulic loading will not change the reaction rate constant. However, a change in the influent BOD will cause a change in the reaction rate constant. Thus, the reaction rate constant is only valid for a particular wastewater at the influent BOD at which the constant was determined.

It is also of interest to evaluate the effect that the second phase kinetic order would have on designs. In the past the phase I kinetic constant has generally been used to express the BOD removal for the entire depth of the biological tower. This may not cause problems if the effluent requirement is not very strict. If a fairly low effluent is required then the use of only the phase I reaction rate constant would provide a design that would not meet the effluent requirement. The phase I reaction rate constant would predict a lower effluent value than that actually achievable.

This evaluation shows that first-order kinetic relationships can be used for designing biological towers if great care is taken in selecting the reaction rate constant.

ORGANIC LOADING RELATIONSHIPS

The total organic loading relationship for designing biological towers has been used with varying degrees of acceptance since the early 1970's. Figure 2 shows typical curves developed from pilot plant data. It is seen that the pulp and paper

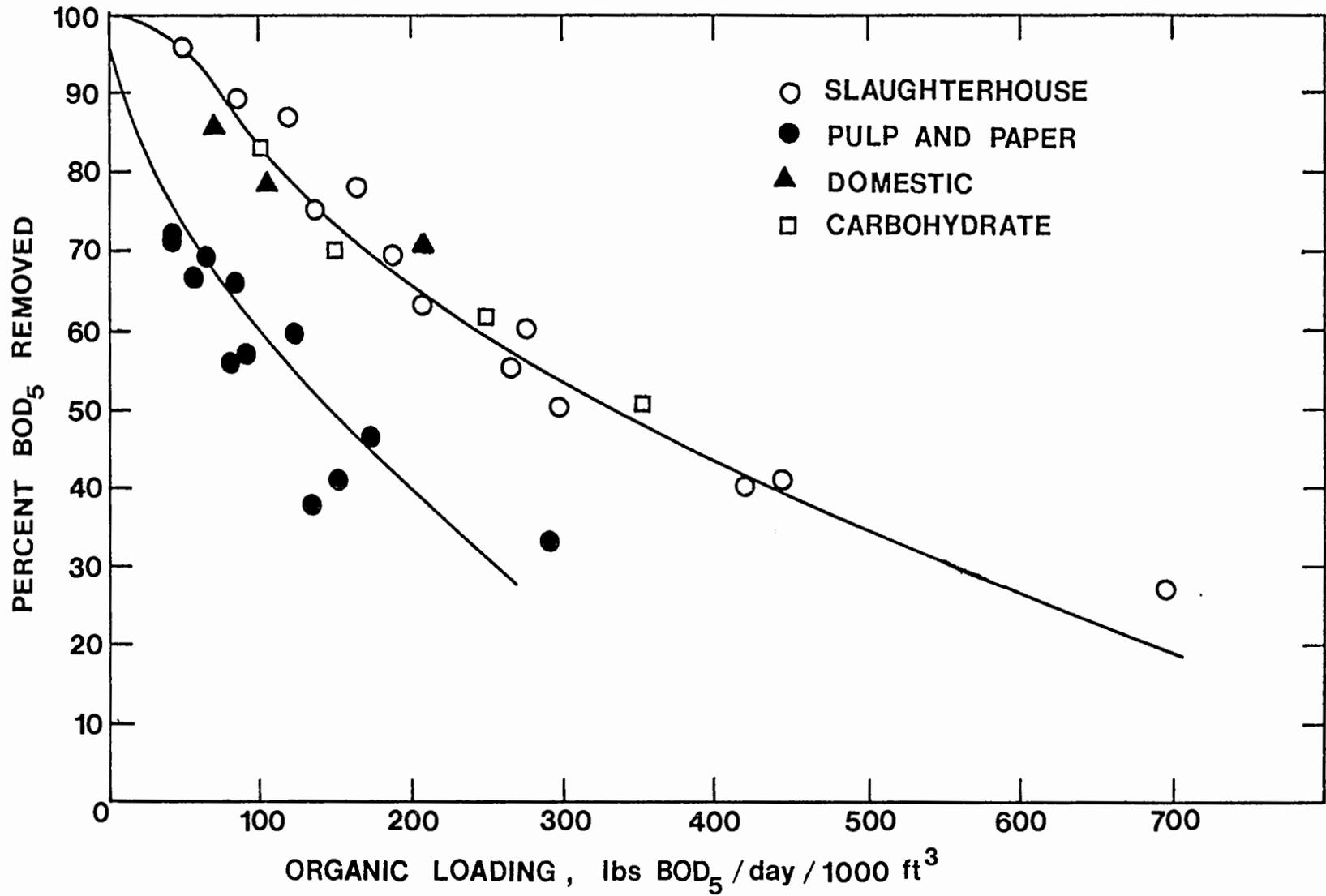


Figure 2. Treatment efficiency as function of organic loading.

wastewater produced one curve; whereas, the slaughterhouse, domestic, and carbohydrate wastewaters produced another curve. This indicates that the relationship is a function of only the type of wastewater and the total organic loading. For a required treatment efficiency, there is an allowable organic loading in lbs BOD/day/1000 ft³. Therefore, a given influent loading will require a given volume of media. It is also important to recognize that the curves developed in Figure 2 are only applicable to media of a given specific area.

A relationship of lbs BOD removed/day/1000 ft³ as a function of the lbs BOD applied/day/1000 ft³ has also been used. This relationship is illustrated in Figure 3. It is seen that the organic removal rate approaches a maximum value. For these wastewaters and media, the maximum removal rate for the slaughterhouse, domestic and carbohydrate wastewaters was 180 lbs BOD/day/1000 ft³ and for the pulp and paper wastewater the maximum removal rate was 100 lbs BOD/day/1000 ft³. The relationship between organic removal and organic loading shows that it goes from a biological rate limiting relationship to an oxygen limiting relationship, thus, explaining zero order kinetics and the saturation concept presented by others.

A relationship similar to that shown in Figure 3 is used for rotating biological contactors. The difference is that the RBC loading factor is per 1000 ft² of surface area rather than 1000 ft³ of media. However, since biological tower media has a specific surface area per cubic foot, the data of Figure 3 can be changed to 1000 ft². This was done for this data by dividing 42 ft²/ft³, the specific surface area of the pilot plant media. These curves are shown in Figure 4. The interesting factor is that the curves shown in Figure 4 are very similar to those found for RBC's (11). The curve for the slaughterhouse, domestic, and carbohydrate wastewaters shows a maximum removal rate of 4.4 lbs BOD/day/1000 ft².

It was also interesting to compare this relationship with field data or pilot plant studies conducted with different media. Figure 5 shows domestic wastewater studies compared with laboratory pilot plant studies. The solid line represents the laboratory studies shown in Figure 4. The symbols represent field and pilot plant studies. It is seen that the field data, which was collected from towers containing media that had a specific surface area of 27 ft²/ft³, compares very well with the laboratory pilot plant which contained media that had a specific area of 42 ft²/ft³. This shows that if organic loadings based upon lbs BOD/day/1000 ft² are used, the specific surface area is not a variable in the relationship. This would

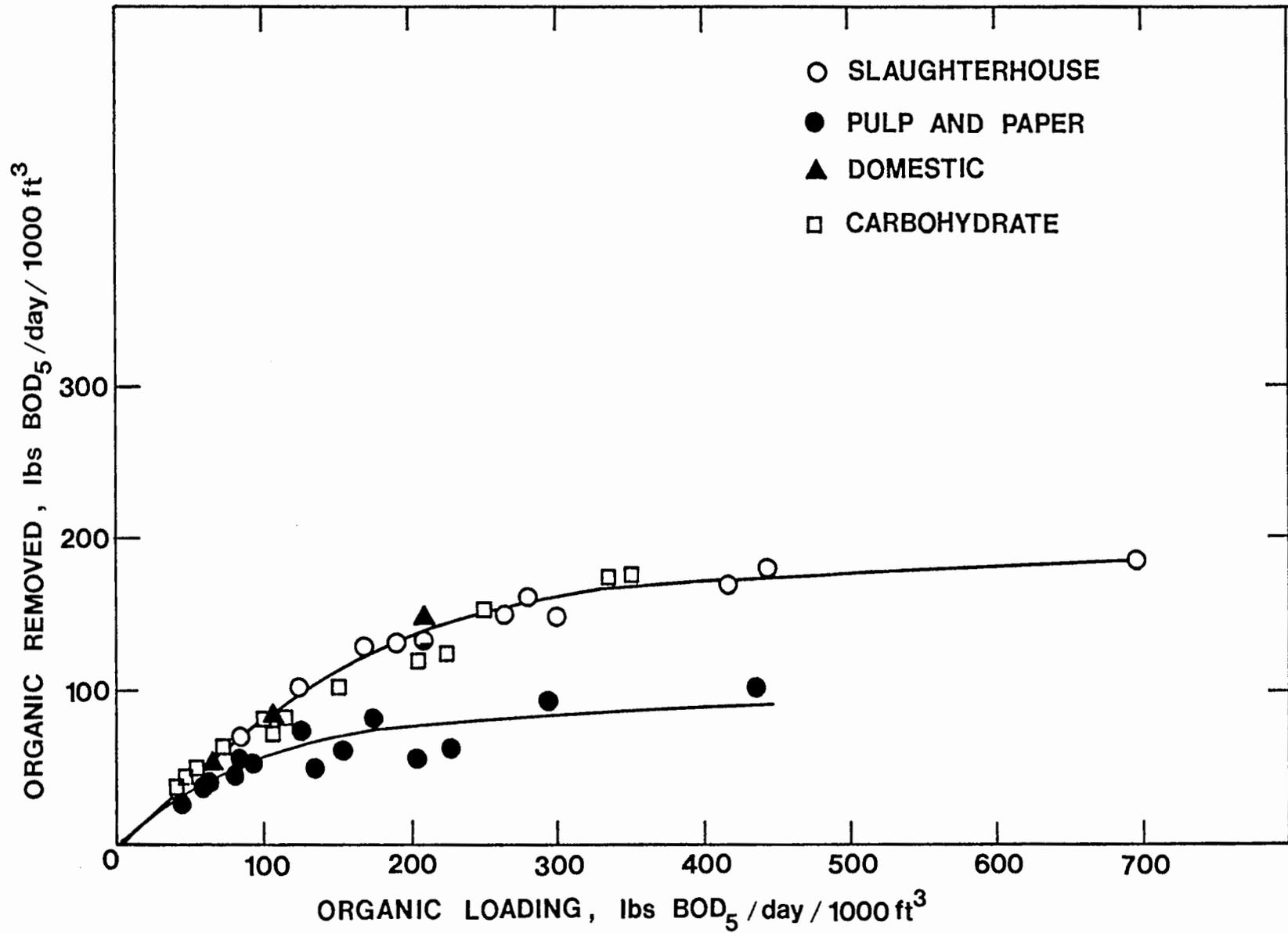


Figure 3. Removal vs. loading (volumetric basis).

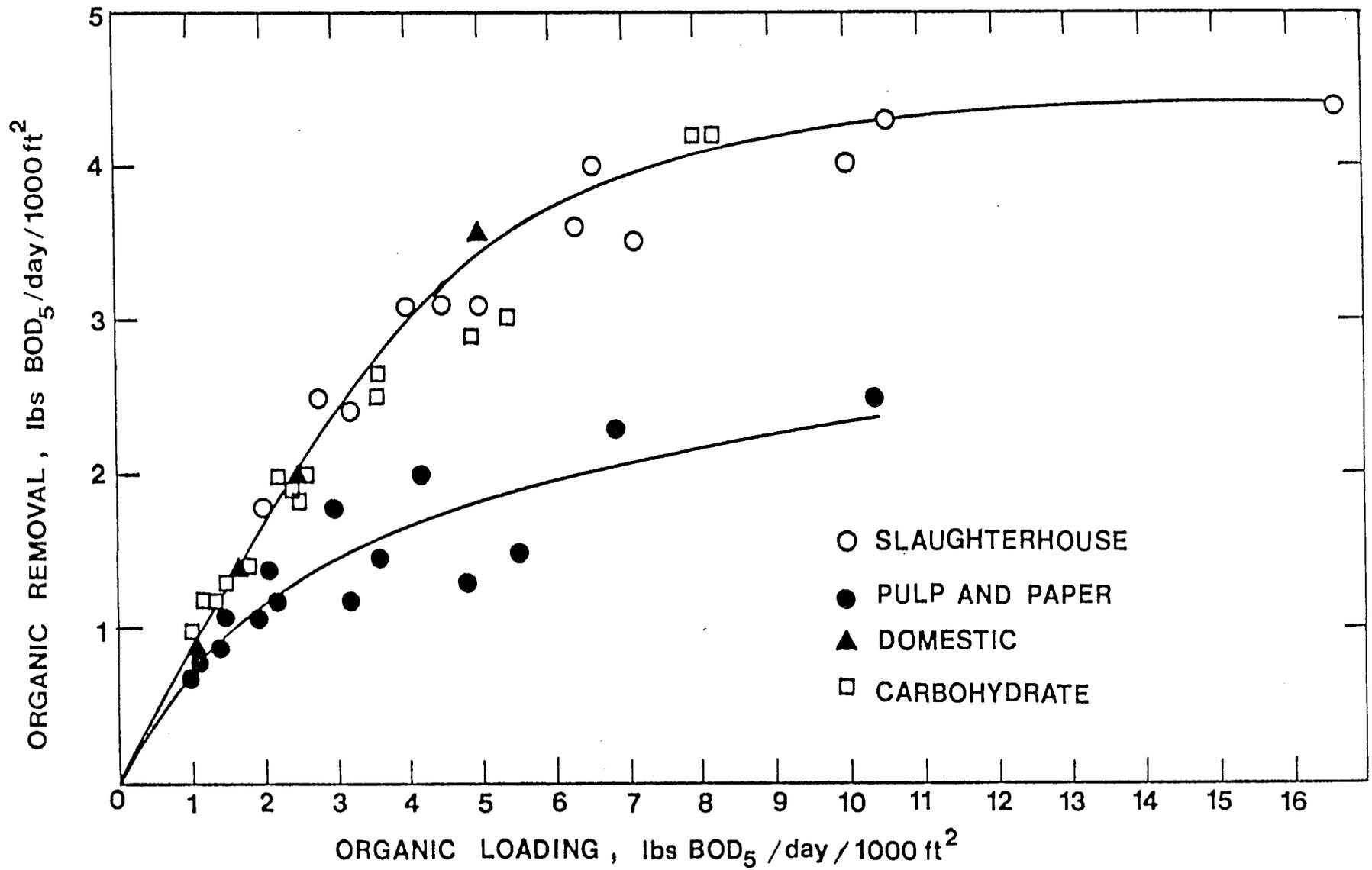


Figure 4. Removal vs. loading (area basis).

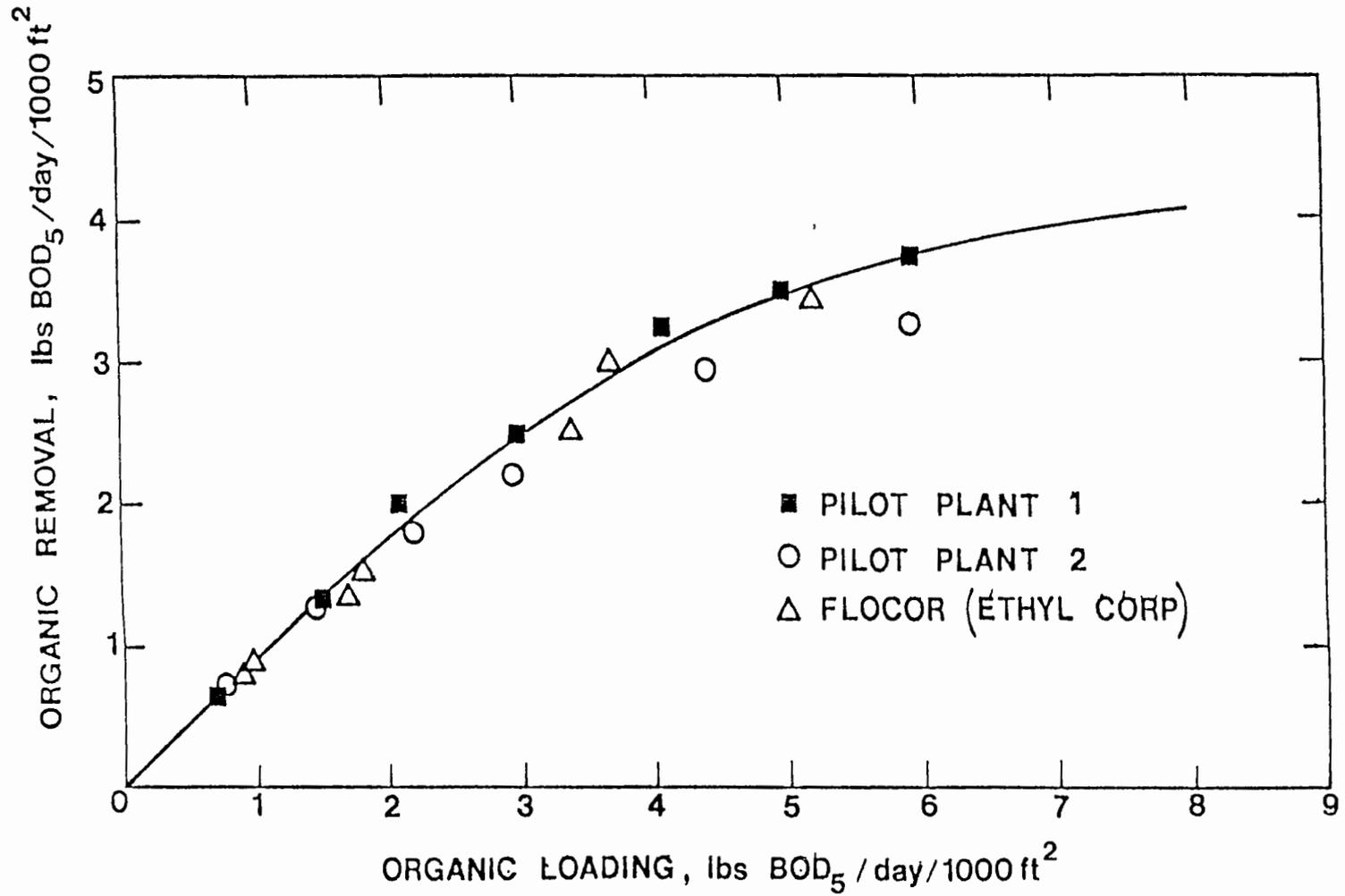


Figure 5. Comparison of field data with laboratory studies.

allow the use of a curve for all media irregardless of the media that was used to develop the curve.

It is also of interest to evaluate what kinetic relationship exists between organic removal and organic loading. It was found that zero order kinetics applied at loadings greater than 5.0 lbs BOD/day/1000 ft². At loadings below 5.0 lbs BOD/day/1000 ft², the kinetics were neither zero order nor first-order. A reciprocal plot of organic removal vs. organic loading is shown in Figure 6. It is seen that loadings less than 5.0 lbs BOD/day/1000 ft² produced a straight line. Loadings greater than 5.0 lbs BOD/day/1000 ft² tailed off from the straight line. This shows that the kinetic order for loadings up to 5.0 lbs BOD/day/1000 ft² follow a "Monod" type relationship. Loadings greater than 5.0 lbs BOD/day/1000 ft² follow zero order kinetics.

TEMPERATURE EFFECTS

It is generally felt that temperature can affect biological tower reaction rates or treatability factors. The accepted practice is to adjust the treatability factor by:

$$K_T = K(20^\circ\text{C})(1.035)^{T-20}$$

$K(20^\circ\text{C})$ = treatability factor determined at 20°C

T = operating temperature, °C

However, our studies show that temperatures above 10°C have no effect on the treatability of a wastewater. This is shown in Table II. This data was collected from a field pilot plant

Table II. Temperature Effects

Temp Range °C	Hydraulic Loading gpm/ft ²	Influent Total BOD ₅ mg/l	Effluent Sol BOD ₅ mg/l	Treatment Efficiency %
22-25	0.75	93	8.6	92.9
18-22	1.25	107	9.5	91.1
14-17	1.25	106	9.5	91.1
10-15	1.25	96	7.5	92.2

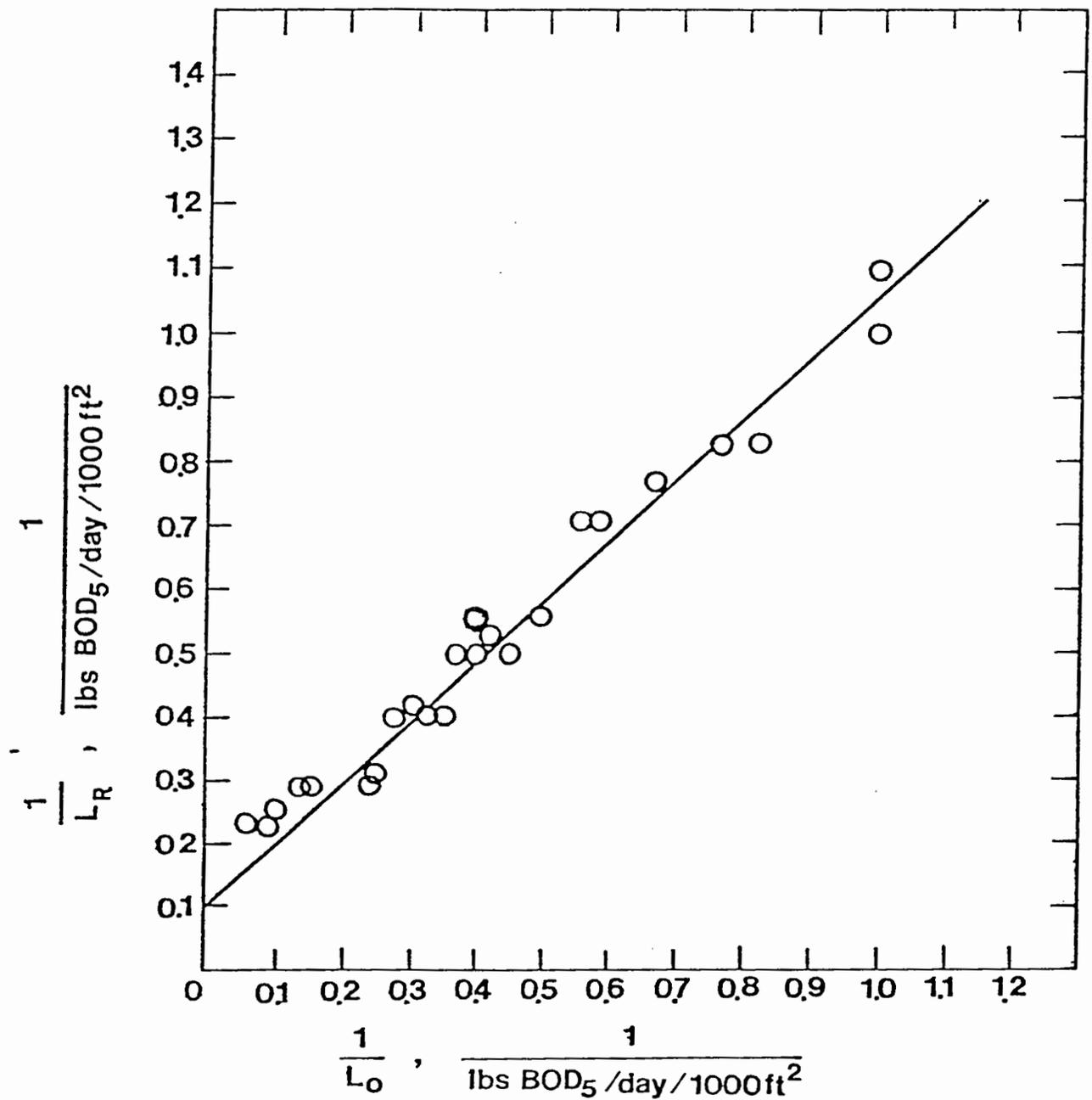


Figure 6. Reciprocal plot of organic removal vs. loading.

treating a domestic wastewater. The temperature varied between 10°C and 25°C during the sampling period. It is seen that the effluent and treatment efficiency achieved at temperatures of 10-15°C are comparable with effluents and treatment efficiencies achieved at the other temperature ranges. Temperatures below 10°C may have an effect on treatability, however, no water temperature below 10°C was recorded during this study. Therefore, no data is available from this study.

RECIRCULATION

A difference of opinion in regards to the effect of recirculation still exists today. Some feel that recirculation must be included in the design equations. However, Germain (12) published an excellent paper in 1966 that showed recirculation has no effect on treatment efficiency. The total organic loading concept also implies no effect. It is felt that this is enough evidence to conclude that recirculation is not a factor in biological tower design.

DISCUSSION AND CONCLUSIONS

The first-order kinetic relationships and total organic loading relationship have been evaluated as design methods for biological towers. The data used in these evaluations were obtained over several years in the Bioenvironmental Engineering Laboratories at Oklahoma State University.

This evaluation has shown that BOD removal with depth of tower does occur as first-order kinetics. However, it has also been shown that at some depth in the tower the removal rate changes and at least two first-order removal rates exist. It was also observed that the removal rate or reaction rate varies with influent BOD, hydraulic loading, and type of wastewater; however, variation due to influent BOD and hydraulic loading can create a serious design problem if this variation is not considered in the design process.

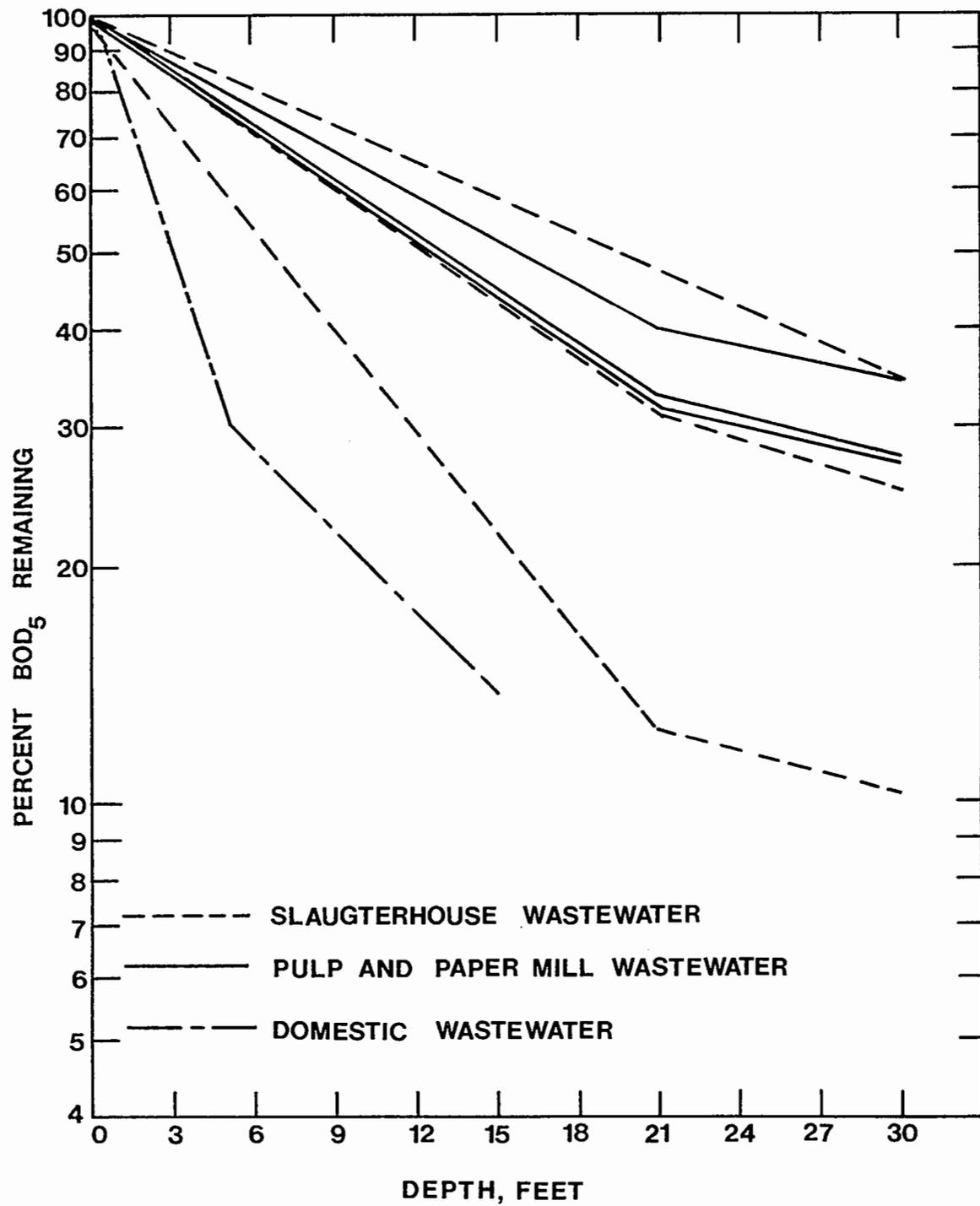
This evaluation has also shown that the total organic loading relationship offers a procedure in which the type of wastewater is the only variation. A design curve can be used with any combination of influent BOD and hydraulic loading. In addition, if a loading based upon square feet rather than

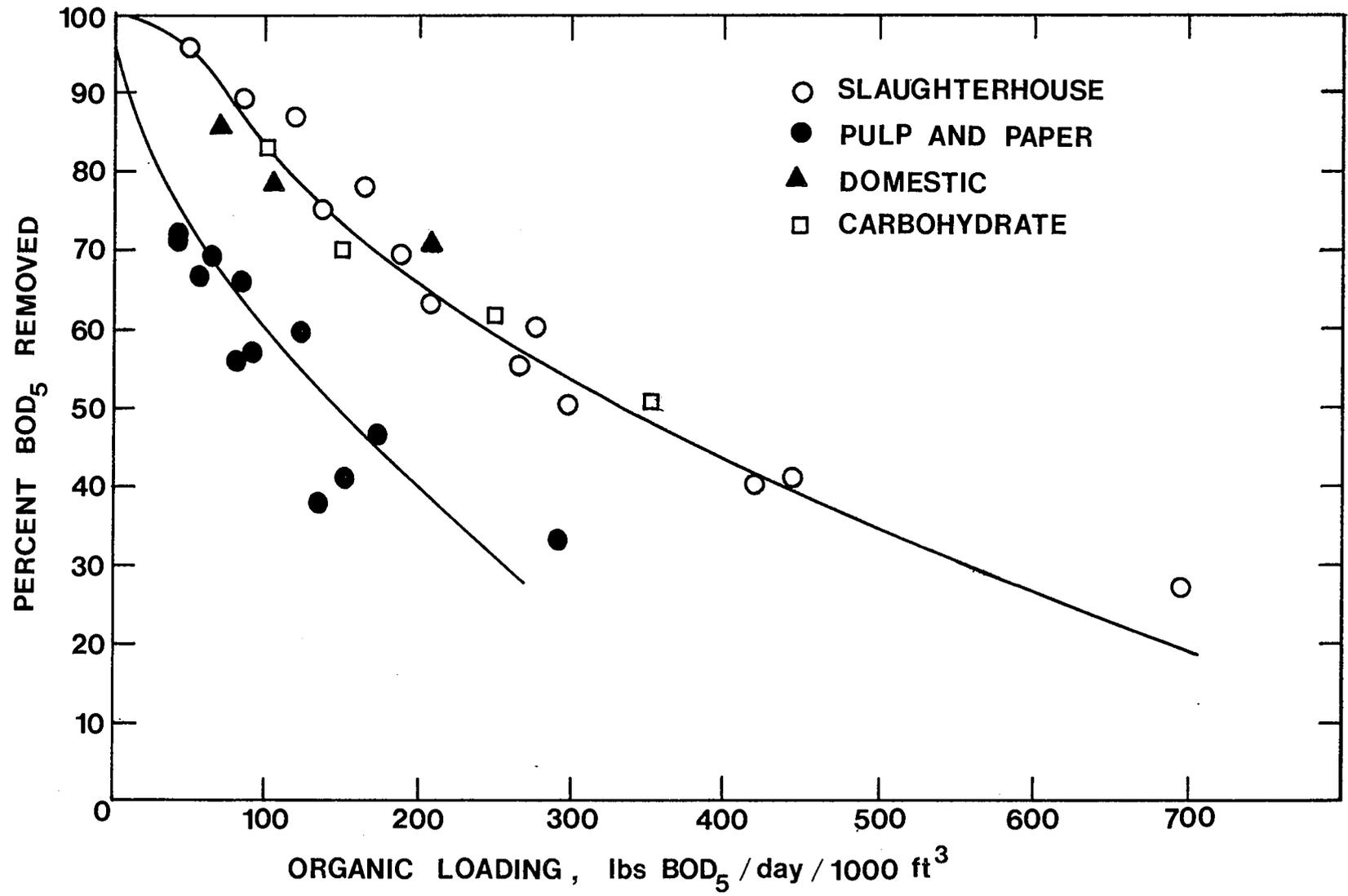
volume is used, the design curve is independent of the type of media.

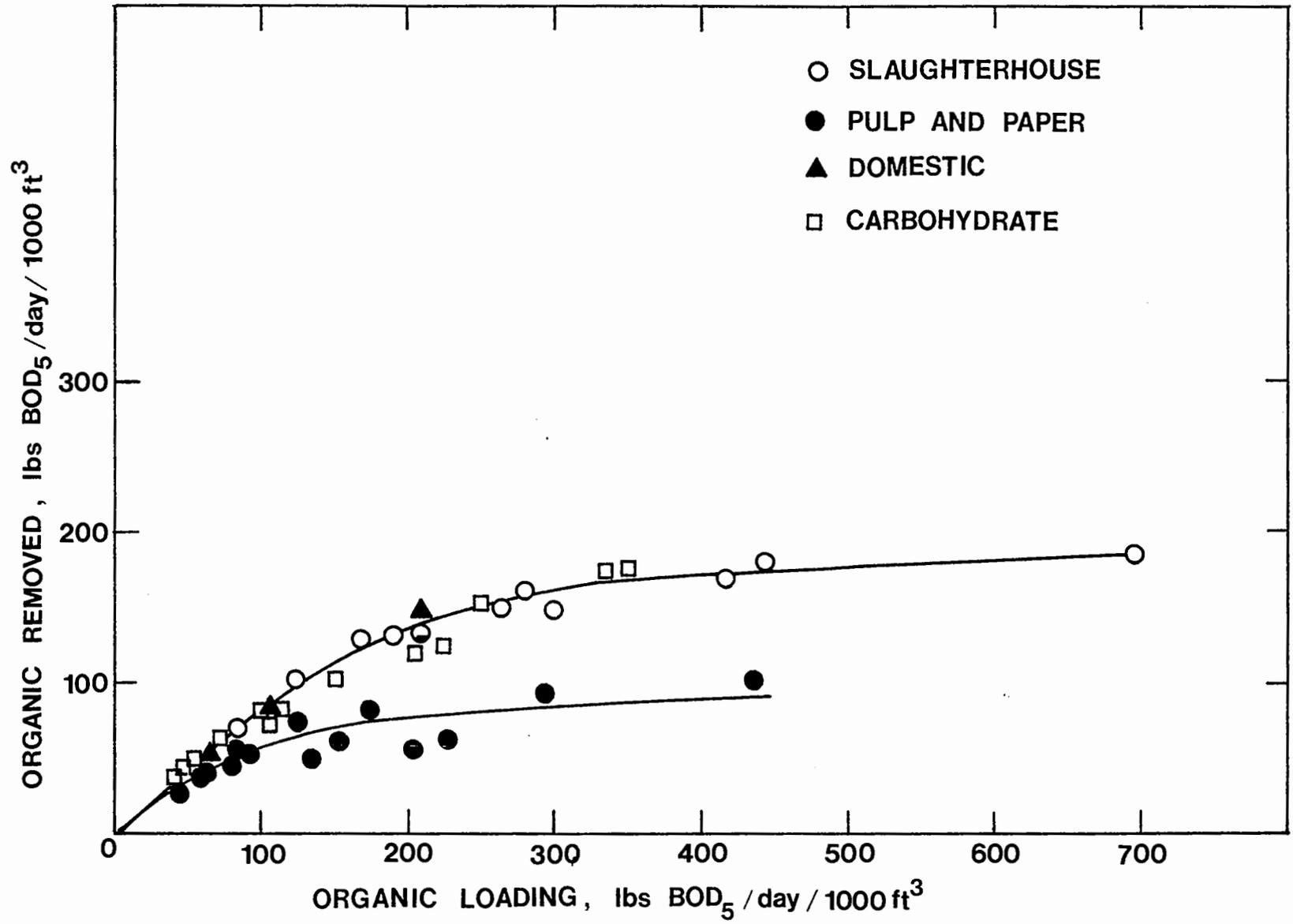
An evaluation of the effects of temperature has shown that temperature variations from 10°C to 25° has no effect on the ability of the biological tower to treat a wastewater.

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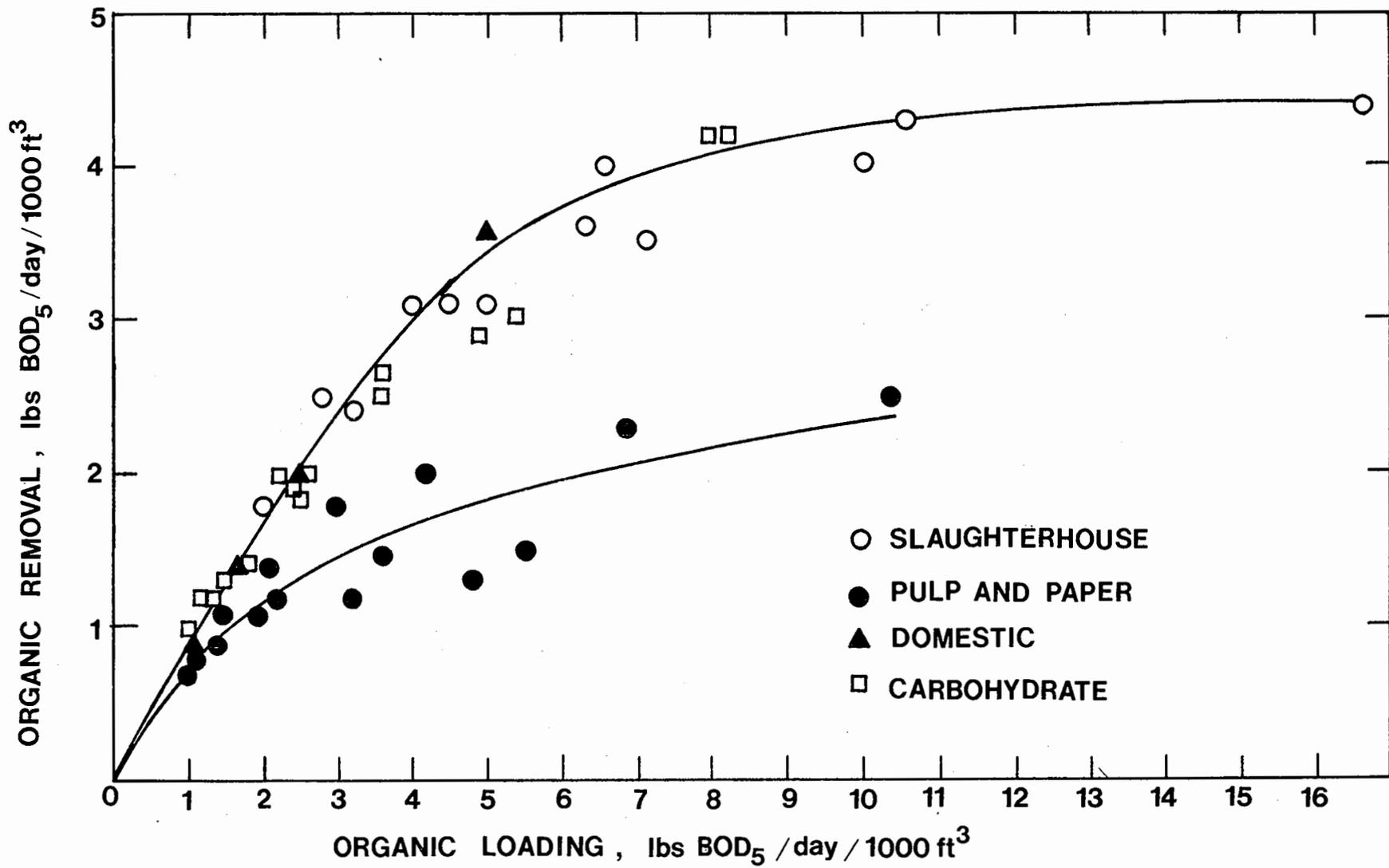
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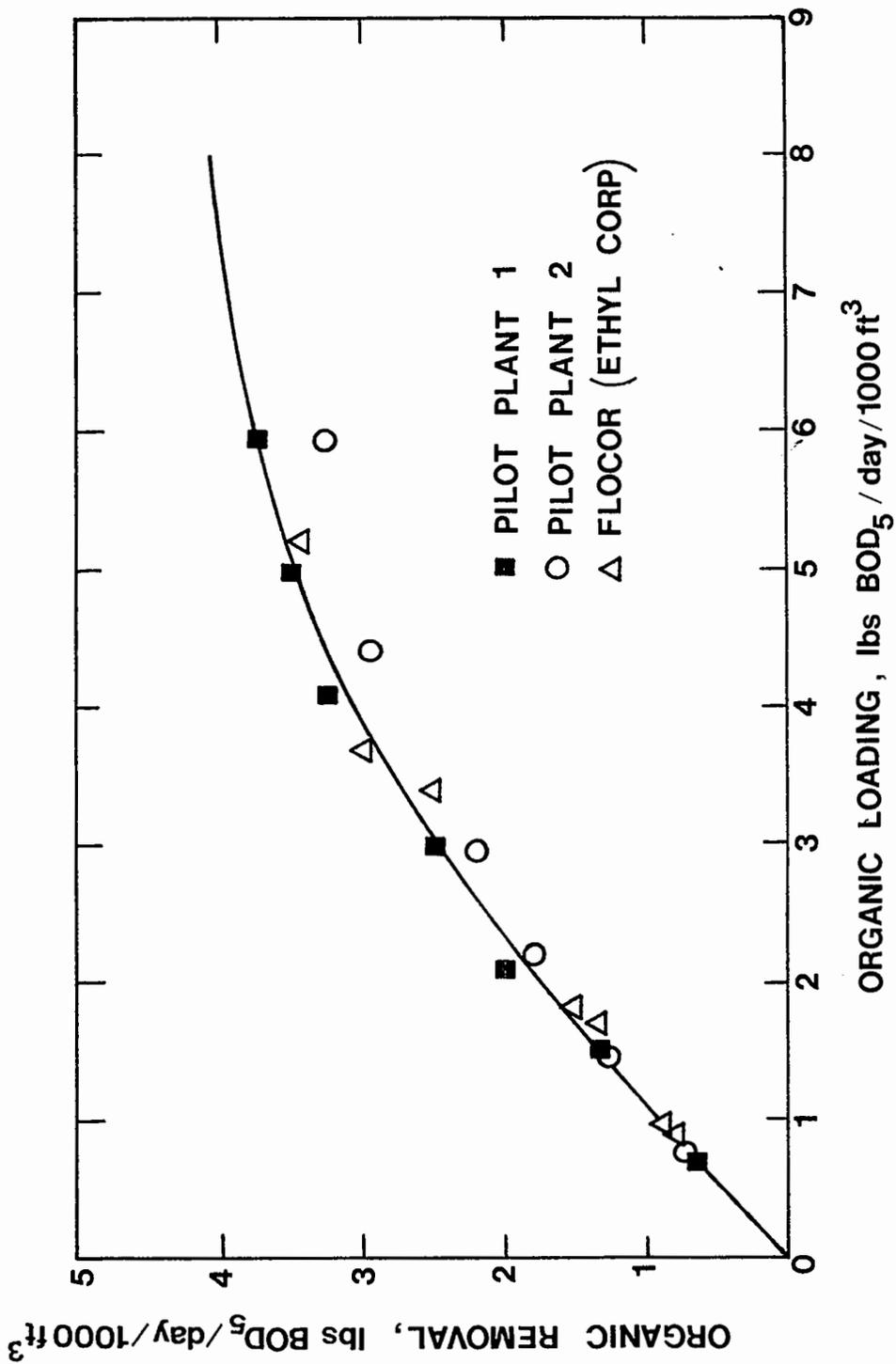


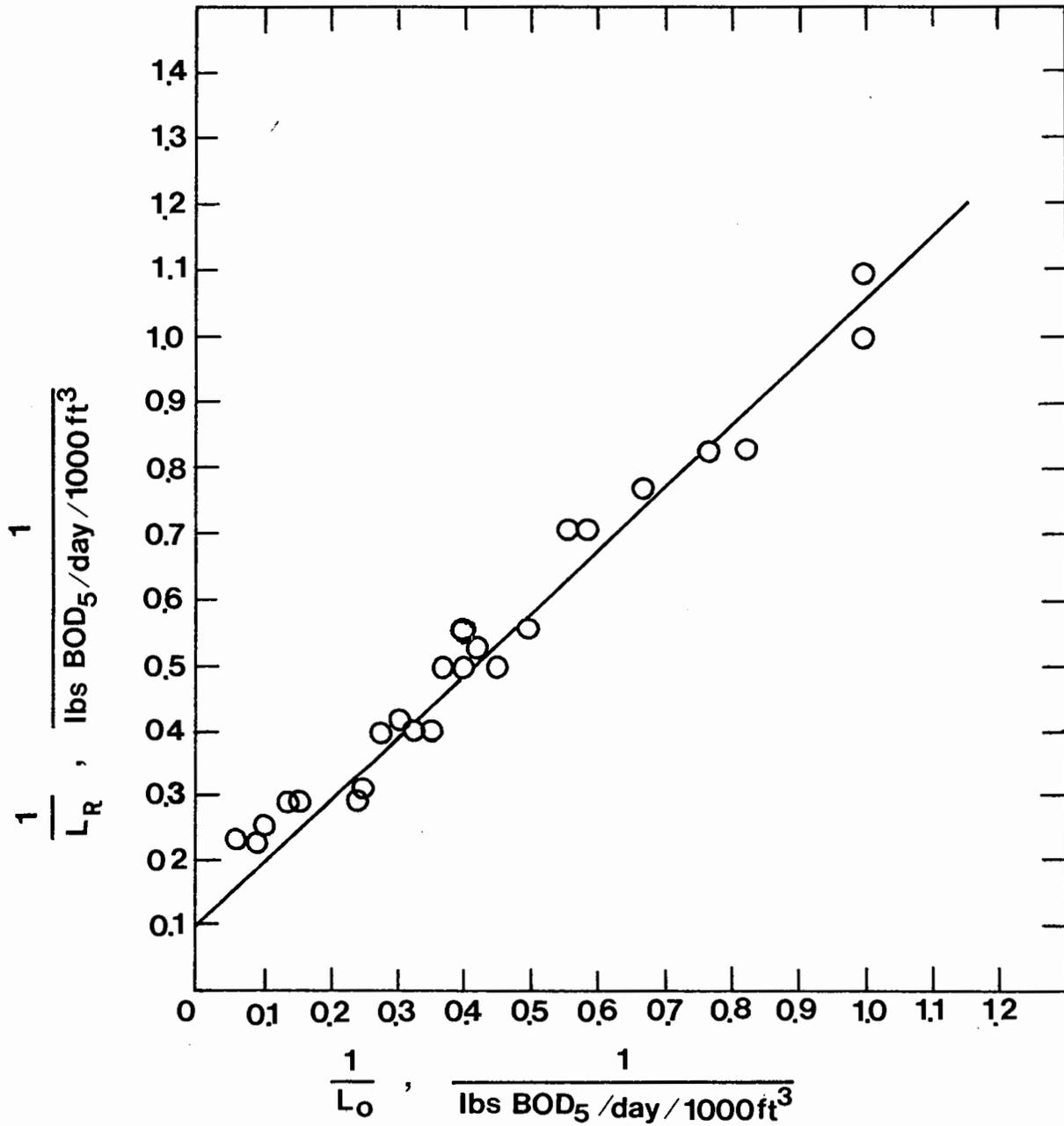




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ANAEROBIC BIOFILTRATION - PROCESS
MODIFICATION AND SYSTEM DESIGN

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INTRODUCTION

Maintenance of long solid retention times (SRT) combined with the proper hydraulic regime - providing adequate shearing of mineralized solids and proper mass transfer conditions - makes the anaerobic biofilter (ANBIOF) the most cost-effective wastewater treatment unit available to a design engineer. With SRT's usually exceeding 50 - 100 days, the ANBIOF can be successfully operated at low temperatures, below the optimum mesophilic range of $35 \pm 2^{\circ}\text{C}$. It is generally observed that long SRTs in ANBIOF reactors promote adequate gasification at lower temperatures, and independence of incoming load changes. Since excess sludge generation is inversely proportional to SRT one can - in certain instances - design a unit producing high quality effluent with almost no excess sludge production.

These features of the process prompted several design modifications of the basic unit originated by Young and McCarty (1). This paper will briefly classify those

modifications, propose the load design approach, present process selection principles and show the need for maintaining significant design flexibility - if the full treatment train is to be the most cost-effective.

AVAILABLE ANBIOF PROCESSES

It appears that the numerous processes available can be classified by the shear exerted on the slime (2). The two groups distinguished are: 1. low shear reactors - with relatively thick slime and low inter-pore turbulence, with practically little mechanical attrition of the slime; 2. high shear reactors - fluidized particulate media covered with dense, thin slime subjected to significant mechanical attrition.

In the schematic in Figure 1, the first group is divided into downflow (e.g. anaerobic RBC or tubular media reactors) and upflow low shear reactors. The second group of ANBIOF reactors may employ inert media - such as sand or reactive media such as activated carbon or an ion exchange resin.

Generally, the first group is characterized by lower permissible loadings - mainly due to less efficient mass transfer and lower slime surface area exposed to substrate. The high shear (fluidized) group 2 is characterized by longer SRT and larger (and more uniform) surface area exposed to substrate/slime mass transfer. Some of the bench scale data on the fluidized bed ANBIOFs indicate such tolerance of the slime to environmental stresses (3) that their performance may be regarded as stable under drastically variable conditions - a state never before achieved in any biological reactor. The current drawbacks of the high shear systems appear to involve the lack of full scale experience, expected high degree of operational skill required and relative sophistication of the process and its control.

DESIGN PRINCIPLES

The list of the design models for fixed film reactors is astounding (4). The complete mechanistic explanation of the processes involved in substrate degradation, under transient conditions of a full scale industrial installation, is still however unavailable. This may be due to the number of factors involved in the mass transfer part

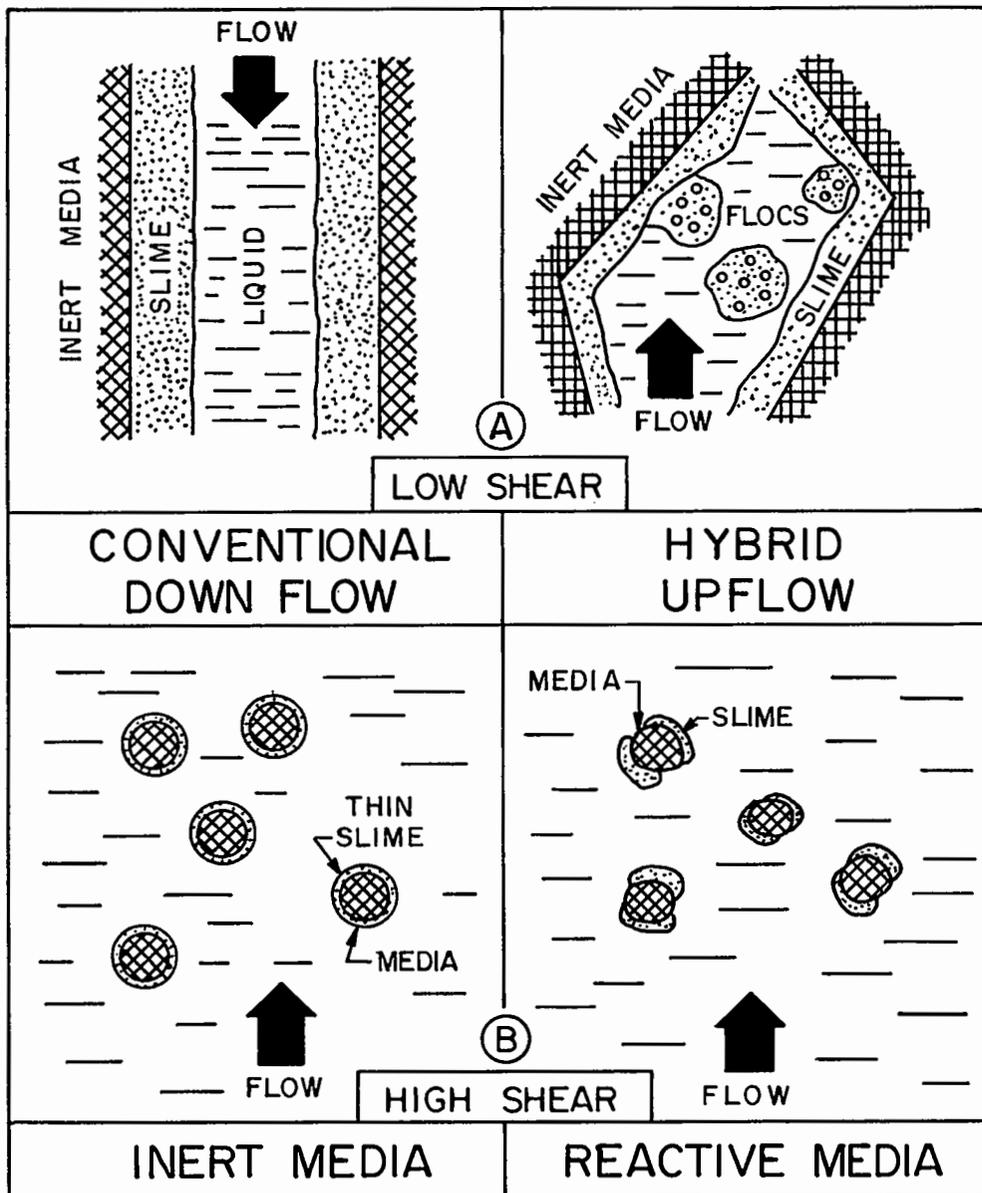


Figure 1 ILLUSTRATION OF DIFFERENCES BETWEEN VARIOUS TYPES OF ANBIOF REACTORS :

- A) Random or oriented stationary media
- B) Expanded or fluidized bed

of the problem and in lack of understanding of the mechanisms of anaerobiosis. It suffices to say that at present at least four distinctive groups (or steps) of bacteria have been identified to be of vital importance in conversion of a complex substrate (e.g. proteins) into CH_4 and CO_2 .

Selection of the Anaerobic Reactor

The decision on the type of an anaerobic reactor is made based on economic comparison of variants (7). This should include a side-by-side analysis of fixed film versus suspended growth reactors followed by an evaluation of the following factors:

- Quantitative variability; e.g. the hydraulic surges are more likely to upset the suspended growth reactor.
- Qualitative variability; e.g. the shut-off mechanism activated in case of toxics is most effective in ANBIOF's (8).
- $\text{COD}_{\text{nf}}/\text{COD}_{\text{f}}$ ratio; high contents of nonfiltered COD suggests the use of suspended growth systems ($\text{COD}_{\text{nf}}/\text{COD}_{\text{f}} > 1.5 \div 2.0$).
- Temperature; lower temperature treatment is much more effective in long SRT reactors such as fluidized ANBIOFs.
- Sludge generation; the economics may dictate containment of sludge in one waste treatment train through application of a suspended growth reactor preceding an ANBIOF.
- Function and place in the treatment train. There are four basic applications: roughing pretreatment; secondary treatment with emphasis on gas production; polishing - denitrification; structure change as in case of color removal for textile effluents.
- Level of expertise available for operation. For small users and intermittent operations, the most "fool-proof" systems are needed.

Outline of the Design Process

The design process consists of the following steps:

- toxicity studies in batch reactors;
- bench scale studies to test loads, gasification

potential and alkalinity requirements in various systems;

- economic analysis and preliminary design;
- large bench scale or pilot scale continuous flow studies testing the depth-load relationship, removal efficiency limits, hydraulic regime, sludge generation and dynamics of process response to influent variability;
- actual design;

Selection of the Design Load

Volumetric organic loading is perhaps the most often used parameter to characterize efficiency of anaerobic reactors (10). The aerobic biofilter under heavy organic loading has been successfully modelled by the load function introduced in 1974 (4):

$$S_e/S_o = \exp(-K/L) \quad (1)$$

where S_e , S_o are respectively effluent and influent COD₃ or BOD concentration; L is the organic loading (kg COD/m³d) and K is the removal coefficient.

Since then the model has been found applicable to anaerobic fixed film reactors (5). Assuming that the slimed specific surface area is equal to A (m²/m³) and that the load is a function $L = Q \cdot S_o/H$ the evolved model may be presented as:

$$S_e/S_o = \exp(-kA H/Q \cdot S_o) \quad (2)$$

where H (m)₃ is depth of the reactor, $K = kA$, Q is hydraulic loading (m³/m²d).

For very high organic loadings, one can further simplify this model using the Maclaurin series expansion:

$$S_e/S_o = \exp(-K/L) = 1 - K/L + (-K/L)^2/2! + \dots \quad (3)$$

which reduces to

$$S_e/S_o = 1 - K/L \quad (4)$$

It is interesting to note that good correlation can be obtained also for the solids retention time (SRT):

$$S_e/S_o = \exp(-K_1 \cdot \text{SRT}) \quad (5)$$

as shown in Figure 2-B. The plot in Figure 2A indicates that two different removal rates apply; K_1 for high loadings, and K_2 for low loadings (below $1.5 \text{ kg/m}^3 \text{ d}$). It is concluded that the low shear upflow hybrid ANBIOF in this example (2) should be loaded above $1.5 \text{ kg/m}^3 \cdot \text{d}$. Lower effluent filtered COD (COD_f) should be attained by employing a second stage reactor.

The use of SRT for design purposes is hindered by difficulties in measuring solids production. The use of the load model (Equation 1) is considered satisfactory for most design estimates. It appears that the high shear fluidized reactors can also be modelled by this equation as shown in Figure 3-A which correlates data reported by Switzenbaum and Danskin (6) for whey waste treatment in a fluidized bed reactor. The data was obtained in varying temperatures $25 - 31^\circ\text{C}$, with hydraulic retention times of 4 to 27 hours.

This performance of a roughing (highly loaded) ANBIOF is followed by an example of an upflow low shear coke-packed ANBIOF polishing an activated sludge effluent - Figure 3-B.

It should be noted that kinetics of COD removal does not always follow the curve with two different dominating removal rates. For simple substrates, the curve may be straight as for synthetic carbohydrate wastes studied by Plummer et al (9) - Figure 4.

Selection of Depth, Media and the Hydraulic Regime

Selection of depth is related to the type of media used and the hydraulic regime applied. The anaerobic biofilter is designed based on volumetric load thus the depth plays a secondary role as far as removal efficiency is concerned. On the other hand, there are certain parameters that determine a depth characteristic for the system: wastewater - media - hydraulic rate. These include: suspended solids distribution, volatile acids distribution, COD concentration profile, level of recycle required for alkalinity control, level of recycle required for maintaining the pseudo-completely mixed system (if needed - as discussed in references 7 and 11).

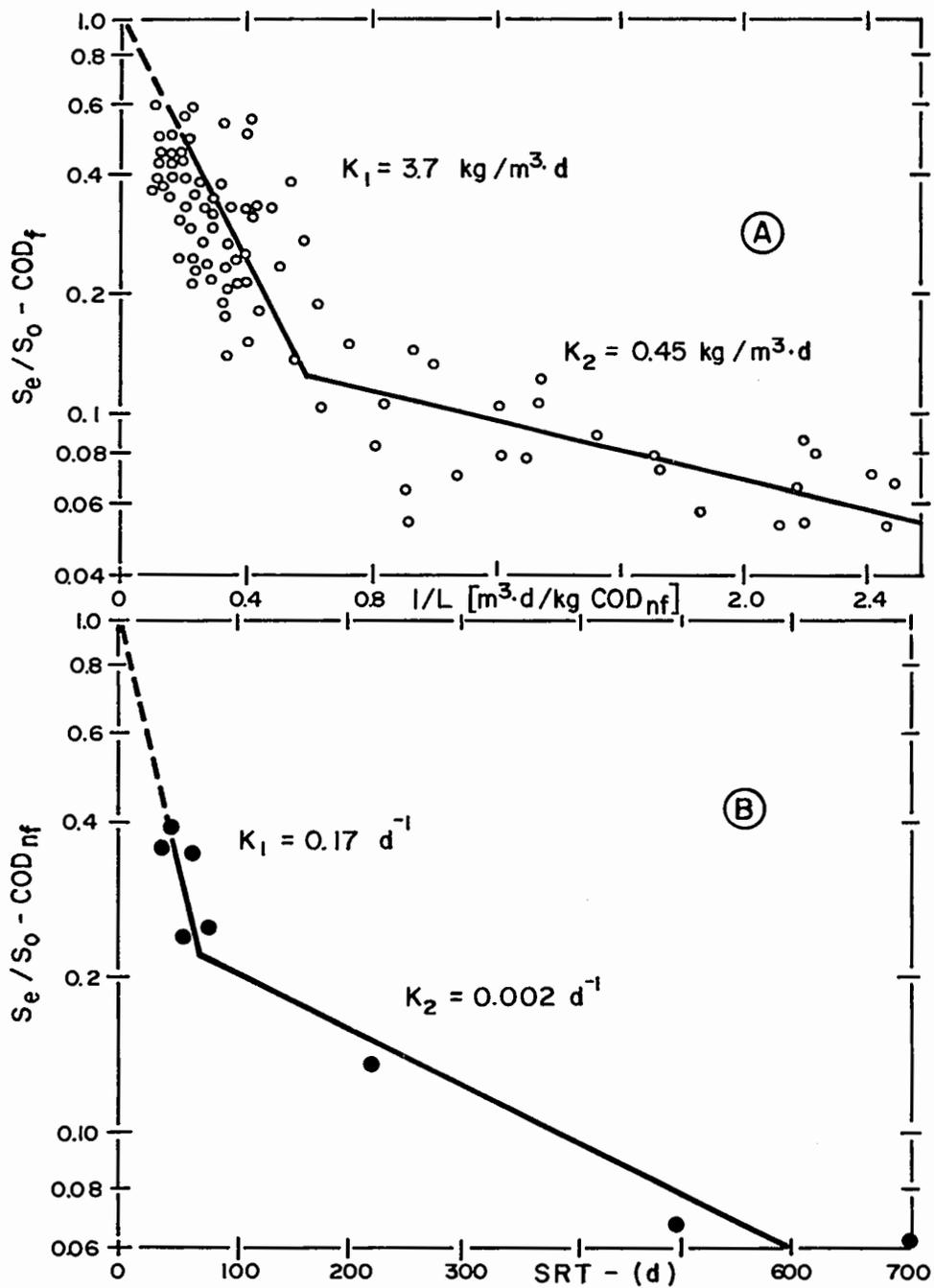


FIGURE 2 COD REMOVAL AGAINST ORGANIC LOADING AND SRT

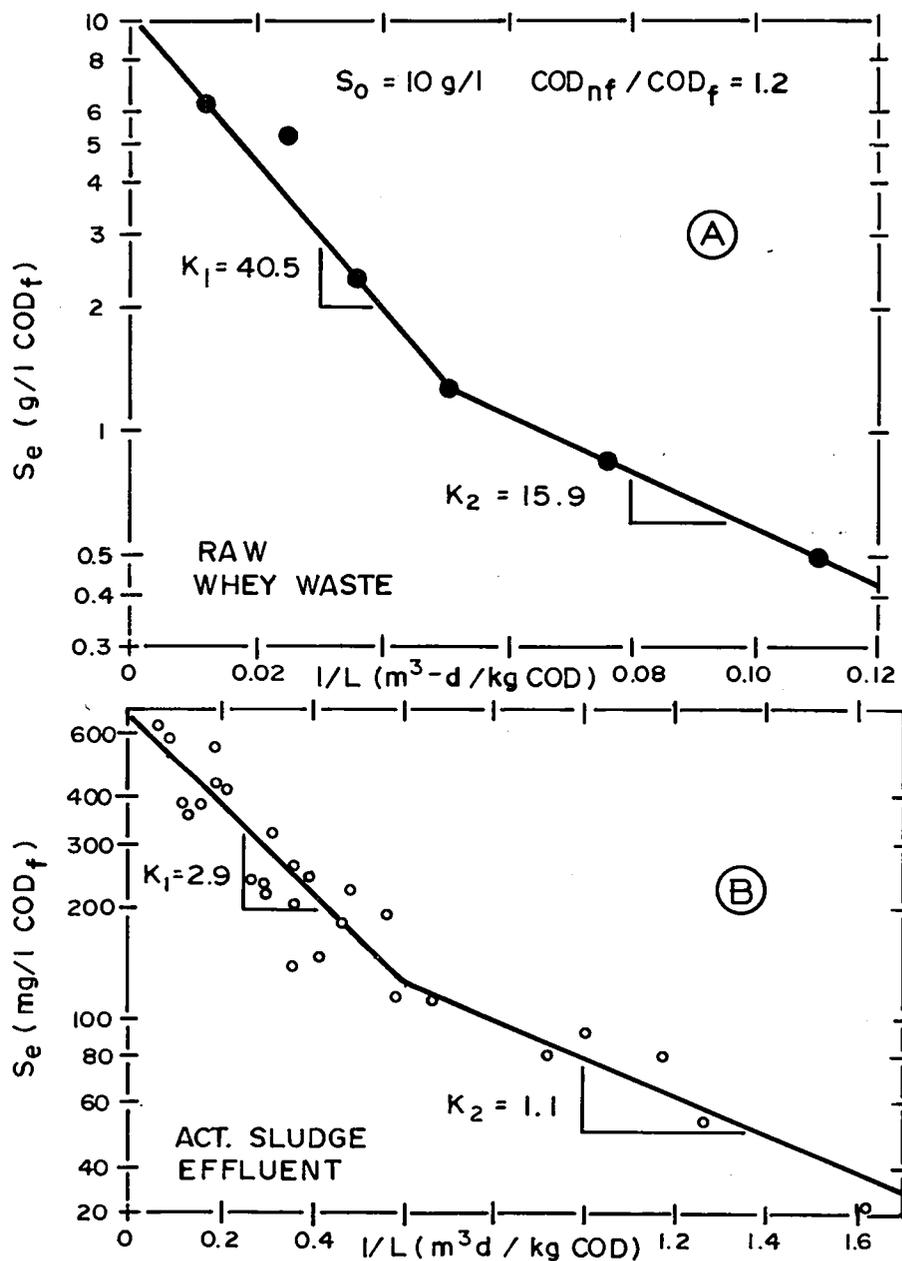


FIGURE 3 KINETICS OF COD REMOVAL:
 A) Fluidized bed
 B) Polishing, coke packed ANBIOF

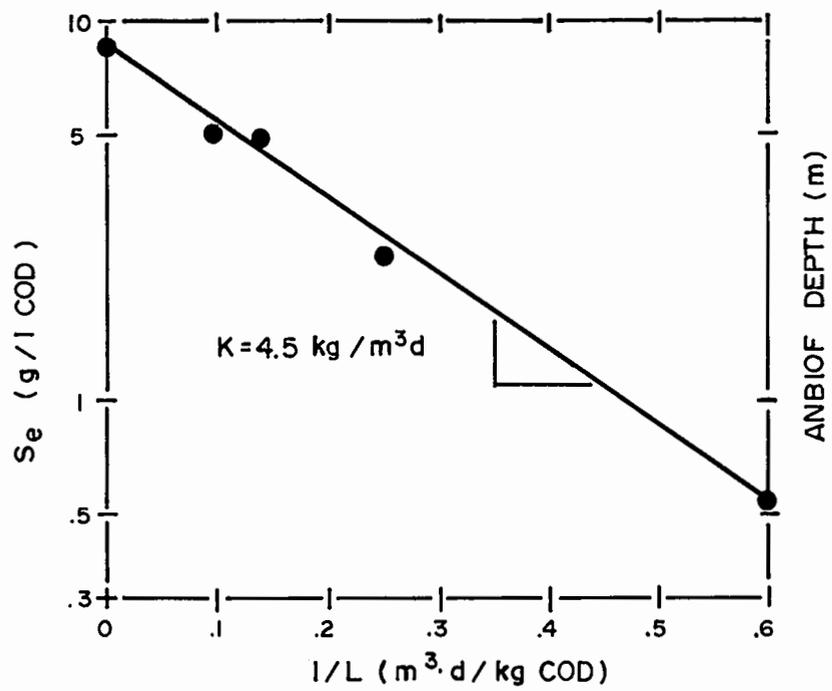


FIGURE 4 KINETICS OF SIMPLE SUBSTRATE DEGRADATION

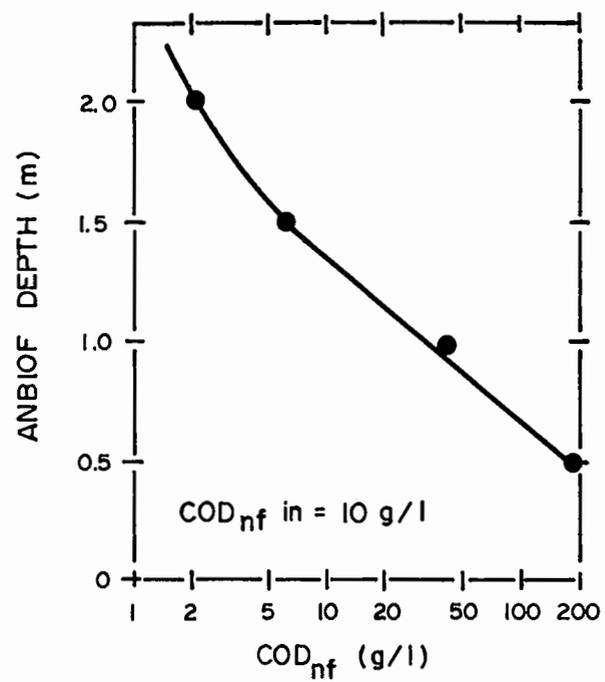


FIGURE 5 COD_{nf} PROFILE IN ANBIOF LOADED WITH SOLIDS

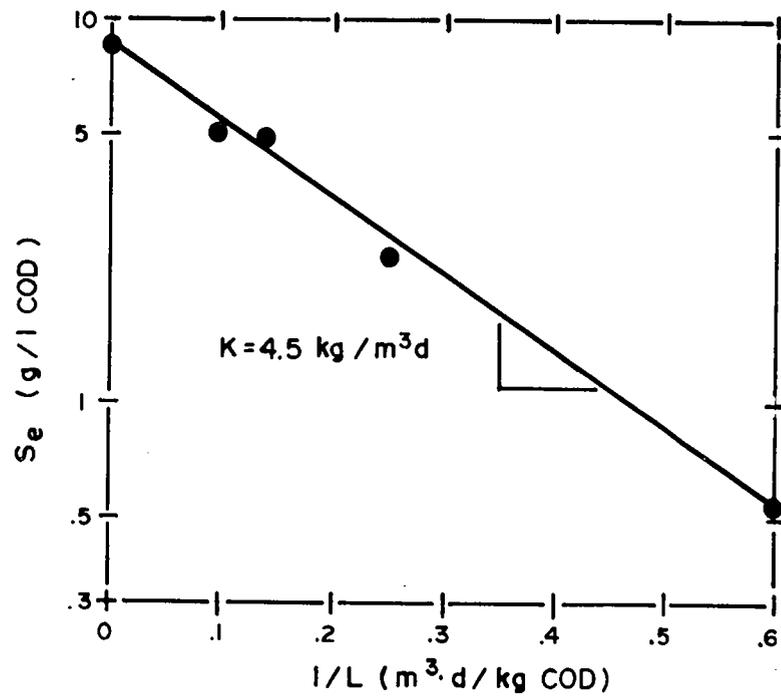


FIGURE 4 KINETICS OF SIMPLE SUBSTRATE DEGRADATION

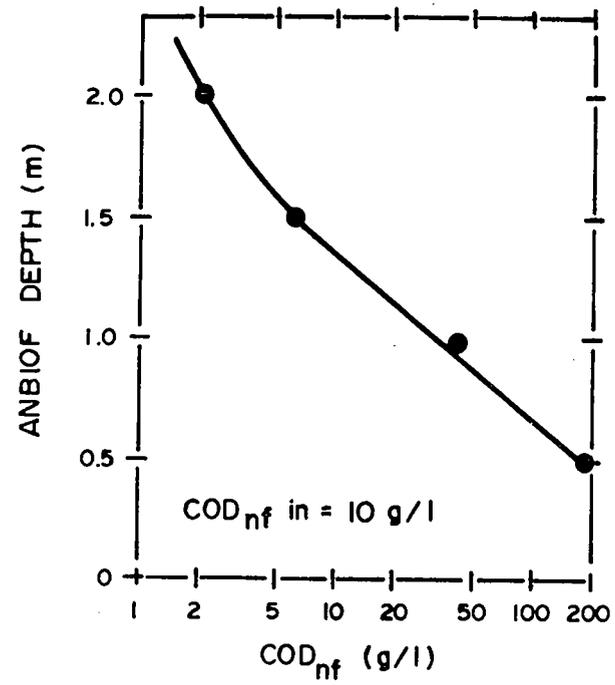


FIGURE 5 COD_{nf} PROFILE IN ANBIOF LOADED WITH SOLIDS

The distribution of solids may be of importance in a porous, stationary media, low shear upflow ANBIOF if considerable amounts of solids are present in the system - and should be retained there. This is illustrated in Figure 5 where solids are expressed as COD_{nf} . The biofilter in this case treated industrial piggery wastes, containing 0.6 - 0.8% TSS and influent COD_{nf} of 10 - 14 g/l ($COD_{nf}/COD_f = 2.0$). The ANBIOF resembled in this case an anaerobic sludge blanket reactor (UASB). The generation of inert solids, beside viable biomass, is also of importance. As found by Johnson & Young (13), the resistance of the anaerobic reactor to toxic substances is improved with the increase of the quantity of inert solids.

The distribution of volatile acids is another important factor in anaerobic biofilters operating without significant recycle. As shown in Figure 6 (taken from Reference 1), full 1.80 m depth is required at the 3.4 kg COD/m^3d loading of this rock filled upflow ANBIOF to achieve adequate utilization of volatile acids.

The depth-organics profile study may reveal cases of excessive depth - at least from the standpoint of solids content and organics removal - as illustrated in Reference 7.

Small scale treatability work should answer most of the problems concerning: effects of temperature, influent variability, gas production, effluent quality, etc. Full scale system design should take into account the differences in the hydraulic regime and the differences in the media used. The design should include also provisions for dealing with inhibitory compounds, toxic slugs and pH upsets.

The removal of organics is dependent primarily on the active volume of the ANBIOF. It appears that the specific surface area of the media in upflow ANBIOFs plays a less important role than in downflow anaerobic biofilters as the overall volume of biomass comes into effect. In scaling-up, care should be exercised to select media with high voids ratio, providing for the most even liquid distribution to minimize channelling opportunities. The configuration of the media should be related to desired upward liquid velocities and the mass of sludge to be kept in suspension (if a hybrid reactor is to be designed).

The oriented plastic media has not been given adequate consideration so far. This is mainly due to apparent difficulties of conducting small scale studies

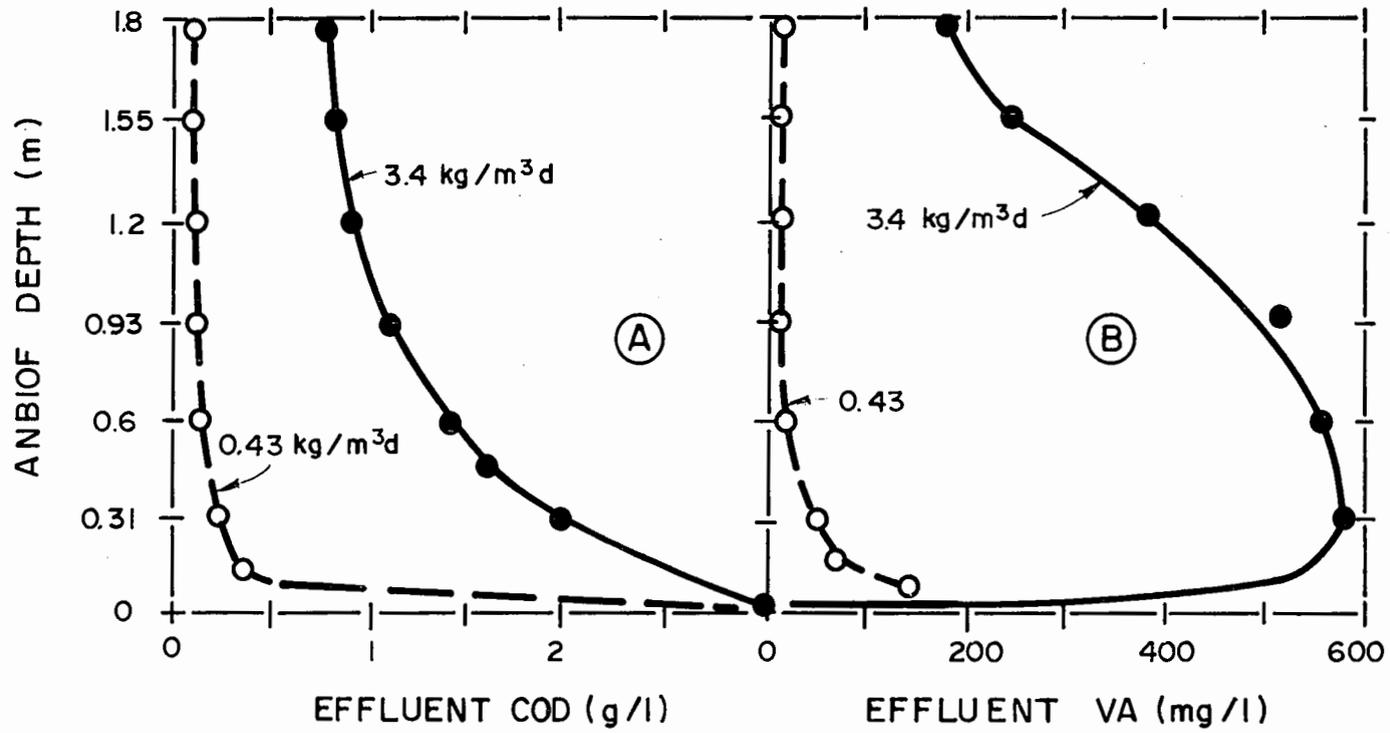


FIGURE 6 ORGANIC PROFILES IN AN UPFLOW ANBIOF:
 A) COD B) VOLATILE ACIDS

with stacked oriented media. Oriented plastic media frequently provides superior hydraulic distribution. The pilot model utilizing oriented media was presented in Reference 2.

APPLICATION AND REMOVALS ATTAINED

The anaerobic fixed film reactors can be applied to all industries generating strong industrial effluents with high content of degradable organics. Most of the full scale installations are within the grain milling, sugar refining, meat packing and fermentation industries. The current trends to concentrate waste streams through in-plant recycle practices offer new challenges to anaerobic treatment technology and often make aerobic systems inapplicable - mainly due to mass transfer limitations of aerobic processes, apparent in conditions of strong wastes at elevated temperatures.

The removals of pollutants are dependent on the loads applied to reactors. The most efficient use of anaerobic biofilters is in the pretreatment mode - before discharge to the municipal sewer. The required COD removals of 80-90% and more are usually achieved without difficulty. The loads recommended to achieve these removals, will depend on the type of waste streams treated, i.e. their relative biodegradability. The list of wastewaters that are presently treated in full, pilot and bench scale is extensive: breweries, distilleries, bottling and soft drink plants, dairies, oil processors, fruit and vegetable processors, canning plants, feed lots, rendering plants, yeasts plants and bakeries, potato processing, pharmaceuticals, pyrolysis effluents, phenolic wastes, pulp and paper, tanneries, and many others. Loads and attained removals for various wastes can be found in references (2, 8, 10, 17).

ECONOMIC EFFICIENCY OF ANAEROBIC BIOFILTERS

In order to break traditional aerobic process selection pattern in favor of anaerobic biofilters, one has to prepare an economic analysis. Size of the facility and complexity of constituents will determine the depth of analysis. For rough comparison, data in Table I is given which compares energy and cost to remove 1000 kg COD in a conventional activated sludge and in course of anaerobic

treatment. It appears that the net savings may approach \$180/1000 kg of COD removed. Thus for 1,550 ton/day corn wet mill, the annual savings over the aerobic treatment could amount to almost \$1.5 million. The costs are based on 1981 prices of \$6.54/ 1000 cu.ft. of gas or \$0.23/m³ (70% CH₄) and 8.0¢/KWh (18), sludge handling and disposal at \$80/ton of solids.

TABLE I
COMPARISON OF PERFORMANCE OF TWO PROCESSES
BOTH REMOVING 1000 kg COD

	<u>Activated Sludge</u>	<u>Anaerobic Biofilter</u>
Power mixing, aeration: (KWh)	1,000	150
Excess biological solids (Kg)	500	50
Sludge conditioning chemicals (Kg)	3	0
Phosphorus requirements (Kg)	10	1
Nitrogen requirements (Kg)	50	5
Cost (\$)*	139.20	17.50
CH ₄ production (m ³)	0	~320
Net (\$)*	(-)139.20	(+)34.50

These savings may be smaller if excess aerobic sludge may be utilized as animal feed or if waste concentration and temperature are low.

Table II shows comparison of capital and operational costs for two systems treating fats and oils processing effluent: 1890 m³/d; S₀ = 6000 mg/l BOD₅. The aerobic train consists of a roughing biofilter₅ followed by a two-stage activated sludge system. The anaerobic train contains a hybrid suspended growth-biofilter reactor followed by a low-shear upflow ANBIOF. The power costs were \$260,000/year and \$15,000 for the aerobic and

TABLE II
COMPARISON OF COSTS OF AEROBIC AND ANAEROBIC TREATMENT
FOR A CPI WASTE STREAM FREE OF SOLIDS

<u>Item</u>	<u>Aerobic System</u>	<u>Anaerobic System</u>
Treatment Units	Roughing Biofilter and 2-Stage Activated	2-Stage: ANHYBRID-ANBIOF
Total Treatment Volume (m ³)	4,300	3,600
Area Occupied by Treatment Units only; m ²	950	610
Capital Cost (\$10000)	1,100	1,100
Excess Sludge Produced (kg/d)	6,300	1,350
Sludge:		
Quality	Requires Anaerobic Digestion or Chemical Conditioning	Ready for Dewatering
Handling (\$/year)	184,000	39,000
Power Requirements (KWh/d)	8,630	500
Power Costs (\$/year)	260,500	15,000
Energy Produced		
Gas (m ³ /d)	0	5,720
Value (\$/year)	0	370,000

TABLE III

COMPARISON OF VARIOUS ANAEROBIC TREATMENT TRAINS
- WASTE STREAM HIGH IN BIODEGRADABLE SOLIDS

System No.	System Description	All Costs Relative to System I	
		14 g COD/l \$0.04/kWh (A)	20 g/l \$0.12/kWh (B)
I	Coagulation - activated sludge - ANBIOF 2° (sludge chemically conditioned)	1.0	1.0
II	ANFLOW (sludge and wastewater) - activated sludge - denitrification	1.24	0.86
III	ANCONT (sludge and wastewater) - ANBIOF - Trickling Filter - ANBIOF 2°	0.85	0.58
IV	Clarifier - ANBIOF - T. Filter - ANBIOF 2° (ANFLOW - sludge)	0.75	0.31

NOTES:

- The costs compared were based on cumulative capital and O&M costs expressed per unit of wastewater volume.
- Both variants are for the same wastewater flow rate.
- ANBIOF 2° denotes denitrifying (anoxic) bio-filter.
- The costs ratio of IB to IA (reference) is 1.37.

anaerobic systems, respectively. On the other hand, the value of recovered gas was \$370,000/year.

Based on these two tables, it appears that in the most favorable conditions, i.e. warm, biodegradable concentrated influent, the anaerobic treatment train may save up to \$170/1000 kg of COD removed.

The comparison analysis changes when significant amounts of settleable, biodegradable solids are present in the raw wastewaters. Then the anaerobic processes efficiency increases with the increase of concentration while they retain their relative independence of solids content. This is illustrated by data in Table III (adapted from Reference 7) which shows the effects of concentration and energy costs on the economic efficiency of treatment trains involving both anaerobic and aerobic reactors. This actual case study has demonstrated that the more concentrated the wastewater and the higher the energy costs - the more economical is the anaerobic treatment technology. The comparison shows also the importance of creative design in combining the anaerobic and aerobic biofilters to arrive at the lowest overall costs index. The use of side ANFLOW (anaerobic flow-through digester) for solids in System IV allows for: 1) use of a series of low volume - high-rate fixed film reactors 2) maintaining a seed material bank 3) utilization of the methane generation potential of all solids generated at the treatment facility. The use of an ANCONT (anaerobic contact digester) reactor followed by ANBIOF may be a preferred route if less costly mixing system is applied. In this case, gas mixing has driven the cost of System III up considerably when compared with System IV.

DISCUSSION AND CONCLUSION

Advantages of anaerobic fixed film reactors over other biological treatment units are:

- No limits set on influent soluble organics concentration.
- No limits set on organic loads applied.
- The lowest sludge (biomass) yield of all available bioreactors.
- The lowest N&P requirements.
- The availability of optional operation in plug-flow and "completely" mixed mode through recycle.

- The lowest level of attention required and associated low operating costs.
- Feasibility of seasonal, 5 days/week or otherwise intermittent operation.
- Highest resistance to toxic and inhibitory compounds.
- Low area (real estate) requirements.
- Odorless operation.
- Generation of high CH₄-concentration (above 70%) gas.

With this impressive, and by no means complete, list of advantages, it is puzzling why the anaerobic biofilter is not used more often. The reasons are: 1) folklore associated with the anaerobic processes and based on poor experience with municipal sludge digestion; 2) limited full scale demonstrations; 3) little understanding of the capabilities of the high rate anaerobic processes; 4) relative ease in coming up with aerobic solutions versus difficulties encountered with anaerobic fermentation.

There are of course numerous unknowns. The recent research, however, indicates that the anaerobic process, if properly acclimated, may be applied to the concentrated industrial organic streams, containing whole array of priority pollutants (8, 12, 13). The anaerobic processes have been shown to degrade compounds which cannot be degraded aerobically (15). From the family of anaerobic reactors, the anaerobic fixed film reactor is singled out to be the most resistant to toxics, shock loads and yielding the highest removals in unstable real-life conditions (11, 14, 16).

The many variations of anaerobic fermenters make the task of selecting the most cost-effective one, difficult. Nevertheless, good technological and economic assessment is needed or the introduction of this technology will be further impeded.

Anaerobic biofilters, due to the long solids retention times (SRT) maintained are the most suitable reactors for soluble COD removal in conditions of variable influent quality and quantity. They are capable of operating at low temperatures and can well withstand slugs of toxic and inhibitory pollutants. The alternative designs of anaerobic biofilters should be carefully screened and tailored to the requirements of the waste stream and the available level of operational assistance.

The volumetric load design appears sufficient for cost estimates. Depth-removal, depth-suspended solids and depth-volatile acids relationships are helpful in selecting the depth and hydraulic regime of the biofilters.

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ROTATING BIOLOGICAL CONTACTOR SCALE-UP AND DESIGN

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INTRODUCTION

The first commercial rotating biological contactor (RBC) was installed in West Germany in 1960. The Allis-Chalmers Company began RBC development work in the United States in the mid 1960's, and there are several companies that presently offer RBC systems for commercial applications. In the United States today, there are more than 150 RBC installations treating municipal wastewaters and over 100 RBC installations treating industrial or special wastewaters. The first of these installations began operations in the 1970's. A historical review of the development of RBC technology has been presented by Dallaire and will not be repeated here (1).

The authors began RBC system process development studies in 1971 and have over ten (10) years experience in treatability, process development and concept design of RBC wastewater treatment systems. All of our studies to date, as well as review of the test results of other investigators and full-scale operations data, have confirmed the applicability of our initial evaluation and design concepts based on the total organic loading approach. This design approach was developed in 1972 and later published (2)(3)(4). This design concept is applicable to design for carbonaceous removal in terms of either BOD, COD, or TOC as well as for nitrification design. This paper presents the culmination of results and our experience from working with the RBC process for the last ten years.

PROCESS DESIGN CONCEPTS (KINETIC CONSIDERATIONS)

Performance data, operating information and kinetic data can be obtained from bench and pilot-scale RBC systems for design of full-scale systems in a manner similar to activated sludge kinetic design approaches. However, there are very important scale-up differences that must be understood and evaluated during the treatability pilot studies for an effective, reliable design relative to performance and economic considerations. The biological mechanisms of wastewater purification are the same for both activated sludge and RBC's; however, the physical differences relative to fixed bed growth versus fluidized bed growth and oxygen transfer characteristics and limitations must be understood and properly considered during the pilot study and scale-up design analysis.

In earlier days, and even today, many engineers involved with RBC's have been concerned with either the hydraulic flow rate or the concentration of substrate (BOD, COD or TOC) applied to the system. The major consideration has been whether the removal efficiency is dependent upon the substrate concentration of the wastewater or the hydraulic flow rate. Some engineers believe that detention time or contact time between the wastewater and the system biota determines the organic removals achievable in the RBC process. Since contact time is related to the hydraulic flow rate, these may be considered to be the same parameter.

Actually, both flow rate and organic concentration exhibit definite relationships with substrate removal rate and efficiency. Our initial studies conducted at varying organic concentrations and hydraulic loadings yielding the same total organic loading in lbs/day/1000 ft² showed that the removal rate and efficiency were indeed dependent on the amount of total organics applied to the RBC rather than its concentration or flow rate (2)(3)(4). The amount of organics removed by the system is the same at the same loading rate regardless of whether the loadings are accomplished by a low flow rate at a high organic concentration or a high flow rate at a low organic concentration. In spite of our initial studies and publications and following publications by others, it is still not uncommon to see the following statement as recent as 1981 in the *Journal Water Pollution Control Federation*. "The question of which parameter, hydraulic loading, or organic loading, to use for proper design and operation of an RBC process has not been resolved."

Others consider the amount of soluble substrate removed per unit of surface area for each stage of a multistage RBC system. This design approach is based on the assumption of first order reaction rate kinetics throughout the multistage system irrespective of initial substrate loadings or effluent requirements. Still another approach considers substrate removal with reactor contact time per stage on the assumption of second order reaction rate kinetics irrespective of initial substrate loadings or effluent requirements. These design approaches do not consider the amount of active biomass on the RBC, the total organic loading applied to the RBC, or the change in reaction rate relative to obtaining high purification efficiency (low effluent BOD's). The gross assumption of first or second order reaction rates can be shown to describe reaction kinetics up to a limiting point; however, design for effluent qualities beyond this limiting point can lead to significant error in scale-up.

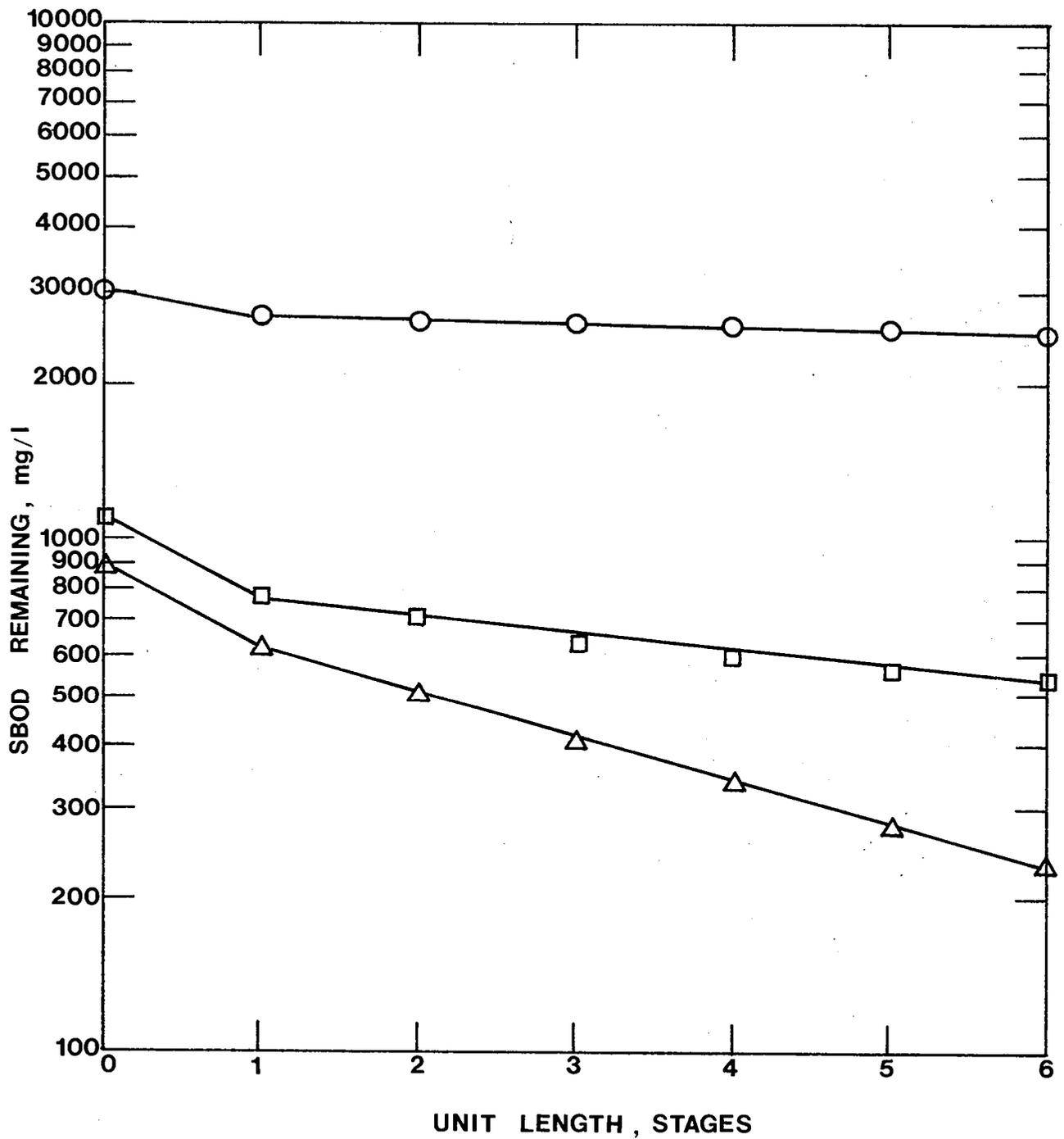
It can be shown in semi-logarithmic plots of substrate remaining versus stages that first order decreasing rates of substrate removal occur in the RBC process. However, in almost every situation with every wastewater there is a distinct slope change following the first stage of the system. These slope changes correspond to a change in the rate of substrate utilization in the following stages. The majority of the substrate is removed in the first stage with the first

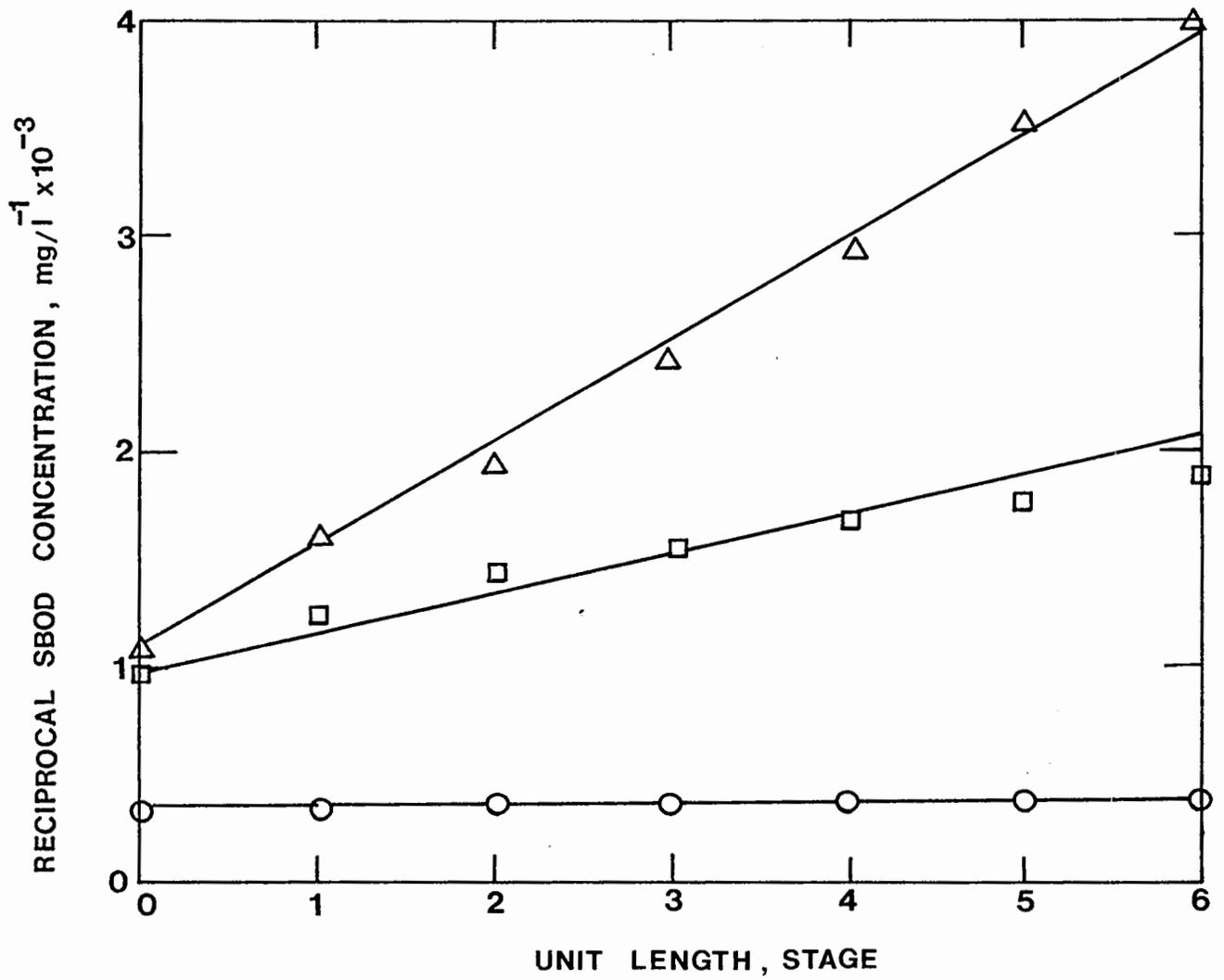
stage removal rate being much higher than the removal rate in the remaining stages. Thus, in order to use a first order kinetic design approach one kinetic constant must be applied to the first stage and a second kinetic constant applied to the remaining stages. However, these substrate removal rates decrease and approach a constant limiting value at the higher substrate loadings indicating saturation with substrate, oxygen limitations or both. At these loading conditions the data show a linear relationship of substrate removal with stage which indicates zero order kinetics.

An example of application of first order reaction kinetics applied to 2.0 foot diameter RBC treating slaughterhouse wastewater is shown in Figure 1. All three BOD concentrations were applied at the same hydraulic flow rate of 0.5 gallons per day per square foot (gpd/ft^2). As previously described there are two separate slopes or kinetic rates for each BOD concentration. The kinetics of the highest BOD concentration appear to approach saturation or zero order kinetics.

The assumption of second order reaction kinetics for substrate removal with reactor contact time per stage has been shown by Opatken (5) to offer advantages over the assumption of first order reaction kinetics. The basic assumptions, except for second order kinetics, are the same as those of the first order design approach. The primary advantage of the second order rate reaction is that one kinetic rate constant can be developed to describe substrate removal throughout the system. As with the first order assumption, this concept cannot predict the substrate removal at and beyond the point of substrate saturation and/or oxygen limitation where zero order kinetics occur. At this point a plot of substrate with contact time shows a linear relationship for the disappearance of substrate.

An example of application of second order reaction kinetics applied to the data in Figure 1, after Opatken, is shown in Figure 2 for the slaughterhouse wastewater where the reciprocal of the BOD concentration is plotted against respective stages. As previously indicated, one slope represents the BOD concentration throughout the reactor, and thus, a second order rate constant can be employed to describe the kinetics. However, as observed in Figure 1, the slope of the highest BOD concentration approaches zero





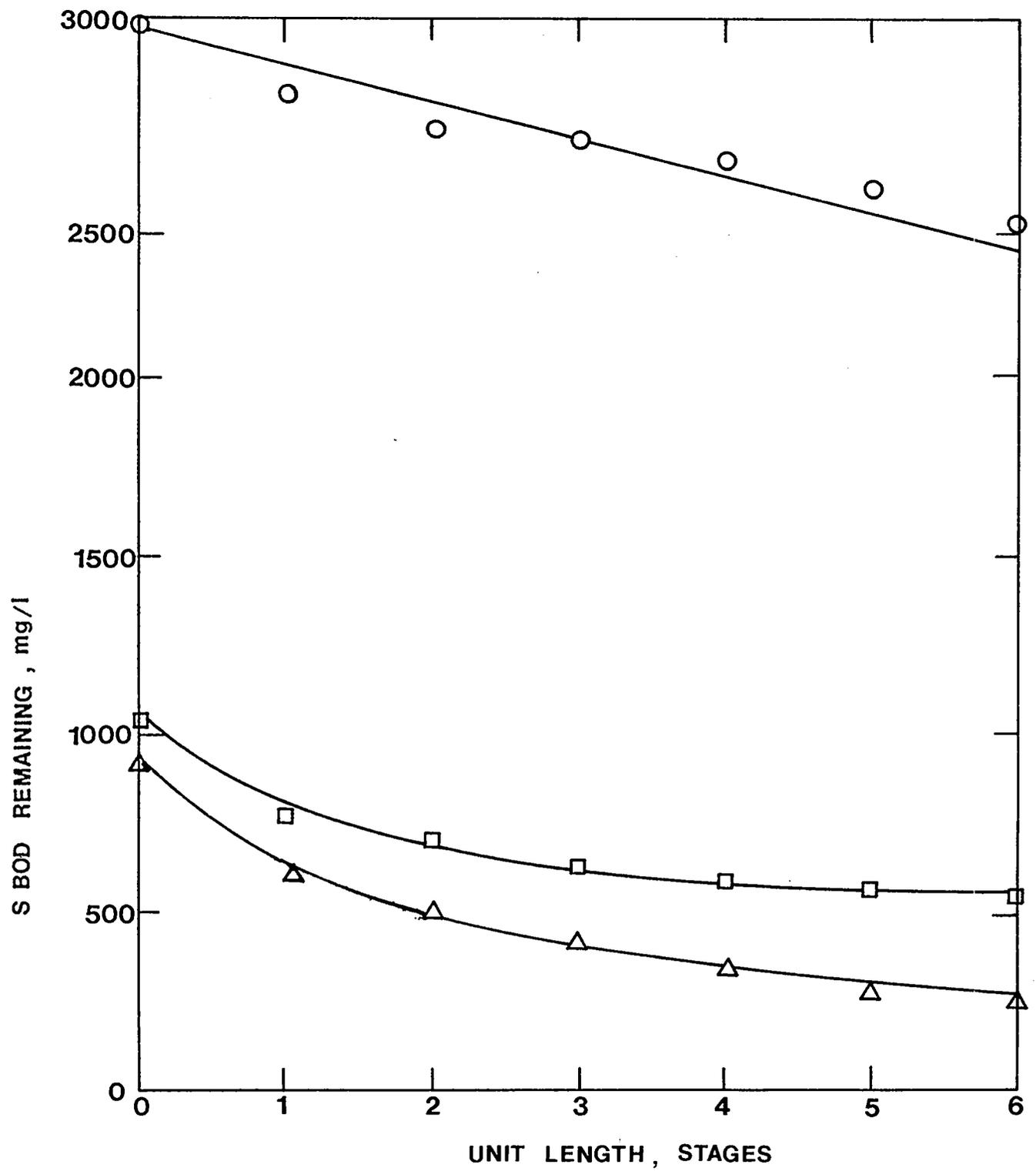
indicating zero order kinetics. A plot of BOD concentration versus stage is shown in Figure 3. A straight line is observed for the highest BOD concentration, indicating that the reaction kinetics for this operating condition are in fact zero order kinetics. The curves for the two lower BOD concentrations indicate that something other than zero order kinetics are occurring. These operating conditions were shown to exhibit apparent second order reaction kinetics in Figure 2.

One advantage of the total organic loading concept is the capability to predict substrate removal and treatment efficiency at any loading condition, irrespective of zero, first, or second order kinetics. Substrate removal relationships are established in terms of the total amount of substrate applied and the loading points or conditions at which zero order kinetics occur can be observed. These points also correspond to loading conditions where the system changes from a biochemical reaction limiting process into an oxygen transfer limiting process. Therefore, the total organic loading concept can be effectively used to determine oxygen transfer capabilities of RBC's, as described in later sections.

PROCESS DESIGN CONSIDERATIONS (OXYGEN REQUIREMENTS)

A major RBC scale-up problem relates to oxygen requirements and oxygen transfer characteristics. A potential oxygen transfer limitation problem exists in full-scale RBC's due to the organic loading on the first stage and oxygen consuming wastewater constituents other than organics, such as sulfides. Oxygen deficiency has been a real problem encountered in full-scale RBC systems due to inadequate pilot studies and scale-up considerations. Exceeding the oxygen transfer capabilities can lead to excessive film thickness and/or proliferation of nuisance organisms which results in a net decrease in organic removal across that stage. A major constraint in the design of any RBC system then becomes limitation of the substrate loading to the first stage to values compatible with the oxygen transfer capabilities of the system.

The design organic loading rate must be evaluated with respect to oxygen requirements, especially in the first

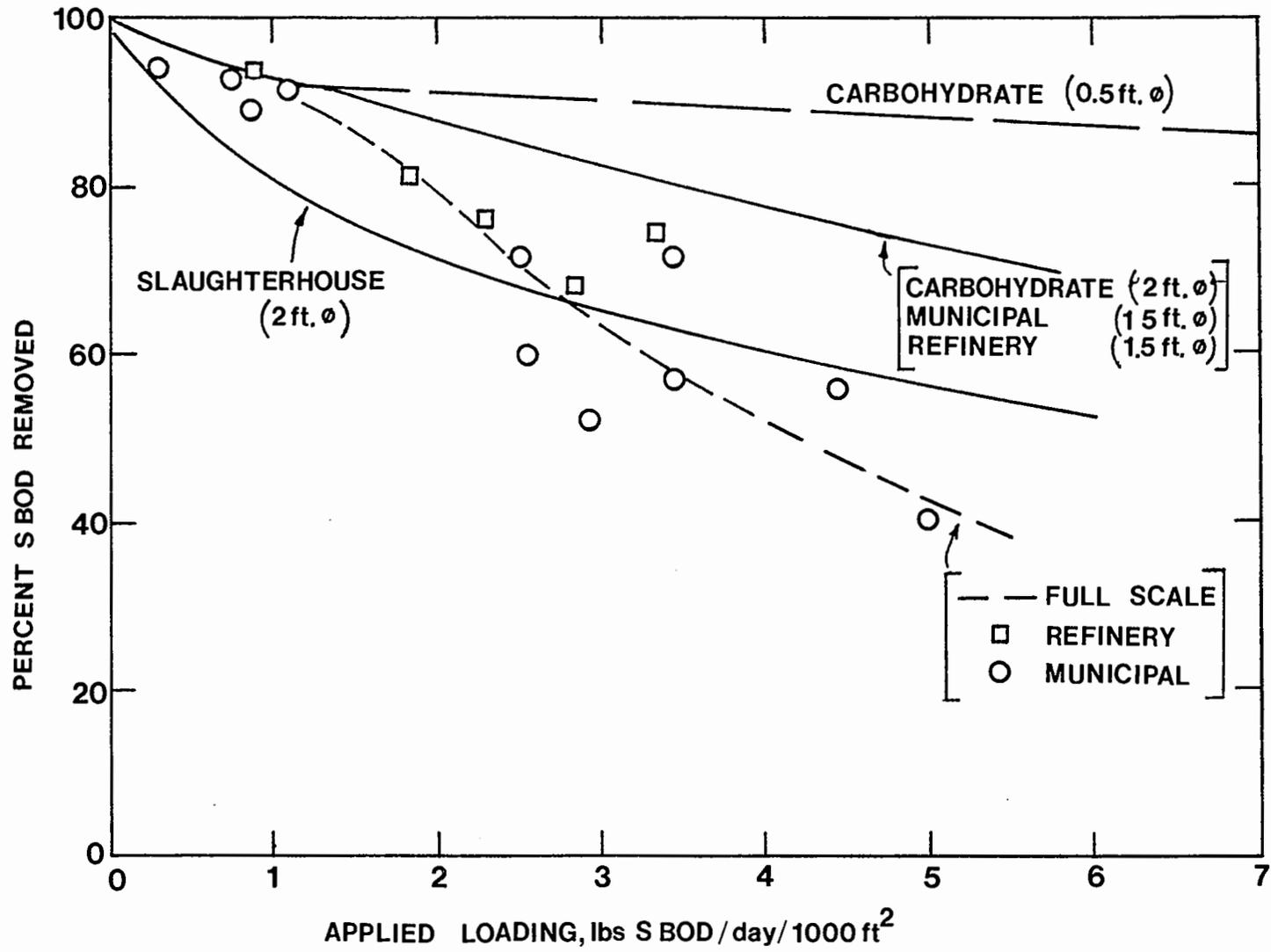


stage. The oxygen transfer characteristics of RBC's varies with the size of the disc and the rotational speed, thus creating problems with direct scale-up modeling for oxygen transfer. The RBC process does not lend itself to the application of standard oxygen transfer tests used for mechanical aeration equipment. Therefore, the oxygen transfer capability of RBC's must be analyzed by other means.

Design parameters, such as disc immersion depth, disc surface area configuration and density, media rotational speed and direction, and surface area to volume ratios may have significant impact on the treatment efficiency of an RBC system; however, these parameters have been standardized for the purpose of optimizing both process design and operation. Current practice requires that about 40 percent of the total disc surface area be submerged in the liquid as a minimum immersion depth. The total effective disc surface area for full-scale installations is determined by disc diameters commonly in the range of 10 to 12 feet. Typical full-scale mechanical drive RBC units are rotated at 1.6 RPM which yields a peripheral velocity of around 60 feet per minute. This rotational speed has been determined by considering that power consumption is exponentially related to the rotating disc velocity. Optimum volume to surface ratios of around 0.12 gal/ft² of media have been reported by several researchers. The surface to volume ratios, geometry and spacial arrangements of each manufacturer's media, rotational speed, and immersion depth have been standardized, and thus the oxygen transfer capabilities of a respective manufacturer's RBC system have been standardized. The problem then becomes one of matching the oxygen requirements with these set oxygen transfer capabilities. The total organic loading concept can be employed in this capacity to match oxygen requirements with oxygen capabilities, as described in the following section.

TOTAL ORGANIC LOADING PROCESS DESIGN APPROACH

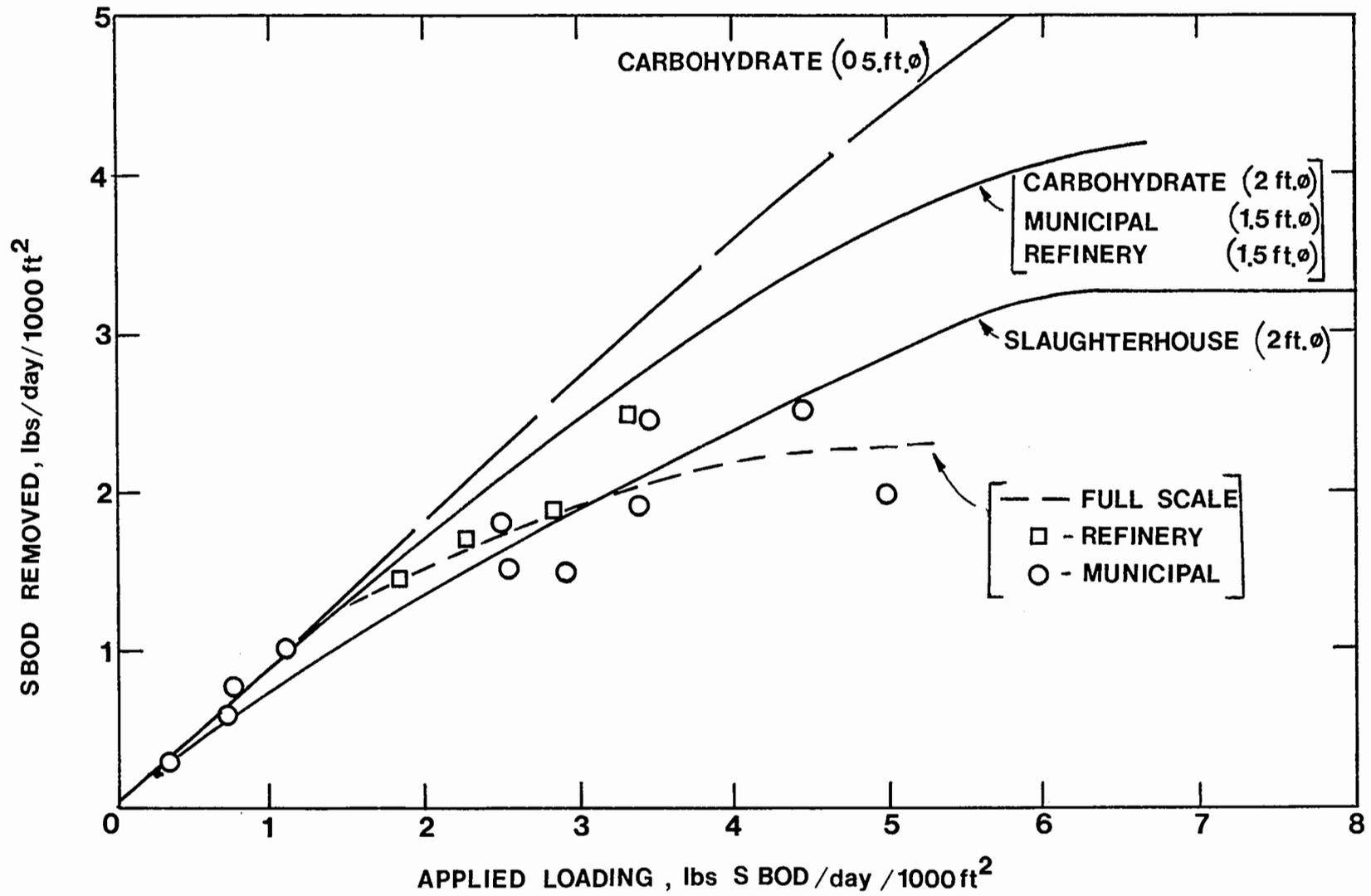
In Figure 4 the combined effect of flow rate and organic concentration, or the total organic loading, on treatment efficiency is shown for several wastewaters with different diameter RBC's. The smaller the diameter of the RBC, the higher the treatment efficiency with the same wastewater. Treatment efficiency of carbohydrate wastewater at organic loadings higher than about 1.5 to 2.0 lb BOD/day/1000 ft²

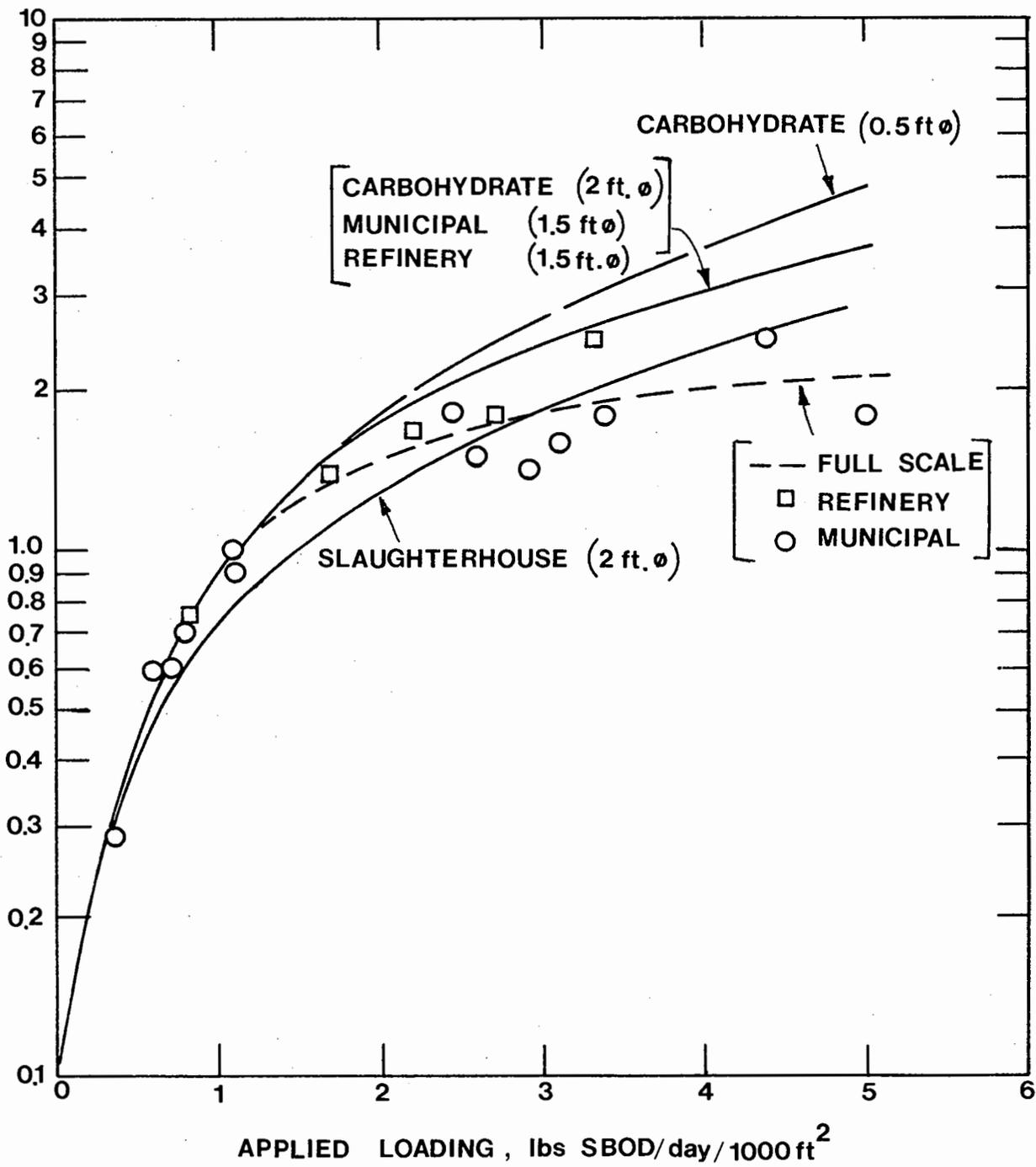


was greater with a 0.5 foot diameter RBC compared to a 2.0 foot diameter RBC. Treatment efficiency of carbohydrate wastewater was greater than the efficiency of slaughterhouse wastewater with the same 2.0 foot diameter RBC at all organic loadings indicating some difference in the biological treatability of these two wastewaters. However, the treatability of carbohydrate, refinery and municipal wastewaters all exhibited the same treatment characteristics. Again, at around 1.5 to 2.0 lbs BOD/day/1000 ft² total loadings, the treatment efficiency of the full scale RBC's started decreasing below that achieved with the 1.5 and 2.0 foot diameter RBC's. These curves show the treatment efficiency of a given RBC treating a particular wastewater at any particular loading rate desired, irrespective of the reaction kinetics (zero, first or second order).

The information shown in Figure 4 can also be translated into a relationship of organic loading removed as a function of the total organic loading applied, as shown in Figures 5 and 6. Figure 5 shows an arithmetic plot of organics removed versus organics applied while Figure 6 shows a semi-log plot of the same relationship. These figures both show the actual organic removal capabilities of a particular size RBC treating a particular wastewater at any desired organic loading rate per 1000 ft² of media surface area. Again, these figures indicate that at organic loadings of around 1.5 to 2.0 lbs BOD/day/1000 ft² the organic removal characteristics of the different diameter RBC changes. The removal efficiencies of the larger diameter RBC's decreases at a greater rate than the smaller RBC's, irrespective of the diameter of the RBC and the wastewater treated. At total organic loadings below 1.0 lb BOD/day/1000 ft² the removal capabilities (lb BOD removed per day per 1000 ft²) of all systems are observed to be essentially the same.

Figures 5 and 6 indicate that the removal capabilities of these RBC's with these wastewaters at the different loading rates did not follow either zero order or first order reaction kinetics in terms of BOD removed per BOD applied. An analysis for second order reaction kinetics also indicated that these relationships did not follow second order kinetics. As shown earlier in Figure 3, the higher organic loading rates can be shown to follow zero order kinetics, and this is also apparent in these figures. Therefore, in these types of kinetic analyses, these curves can be shown to follow





different orders of reaction kinetics as the organic loadings are increased. The breaking point of 1.5 to 2.0 lbs BOD/day/1000 ft² for all RBC's, where the amount of BOD removed per BOD applied begins to significantly decrease, is more readily apparent in Figure 6 compared to Figure 5. At and beyond these loading conditions the removal capabilities decrease in all systems and the scale-up differences due to different diameter RBC's can be readily observed.

Table I presents the difference in treatment capabilities of the 1.5 foot diameter RBC's compared to the full scale RBC's when treating refinery and municipal wastewaters, as determined from the curves in Figures 5 and 6. Table I shows the BOD removal capabilities of these systems between 1.0 to 4.5 lbs BOD/day/1000 ft² total loadings. Up to 1.0 lb BOD/day/1000 ft² the removal capabilities of the full scale RBC and the 1.5 foot RBC are the same, and the removal capabilities do not differ significantly until the total loading is increased beyond 1.5 lb BOD/day/1000 ft².

Table I. Pilot Versus Full Scale RBC Treatment of Refinery and Municipal Wastewater

S BOD Loading lb/day/1000 ft ²	S BOD Removed, lb/day/1000 ft ² (Percent Removed)			
	Pilot Scale (1.5 ft)		Full Scale	
1.0	0.92	(92.0)	0.92	(92.0)
1.5	1.30	(86.7)	1.25	(83.3)
2.0	1.70	(85.0)	1.50	(75.0)
2.5	2.10	(84.0)	1.70	(68.0)
3.0	2.50	(83.3)	1.80	(60.0)
3.5	2.80	(80.0)	1.90	(54.3)
4.0	3.10	(77.5)	2.00	(50.0)
4.5	3.40	(75.5)	2.00	(44.4)

An apparent dependence of the BOD removed on the BOD applied is shown in Figures 5 and 6. It is seen that as the BOD applied approaches higher and higher values, the BOD removed approaches a maximum value where further increases in BOD applied cause no further increase in BOD removed

(zero order kinetics). The relationships shown in Figures 5 and 6 can be satisfactorily fitted with a hyperbolic function similar to the "Monod equation" as follows:

$$L_R = \frac{L_{R(\max)} L_0}{K_S + L_0} \quad (1)$$

where

L_0 = Applied BOD loading in lbs BOD/day/1000 ft²

L_R = BOD removed in lbs BOD/day/1000 ft²

$L_{R(\max)}$ = Maximum BOD removed in lbs BOD/day/1000 ft²

K_S = Applied BOD loading rate at which the rate of BOD removal is one-half the maximum rate, or the saturation constant.

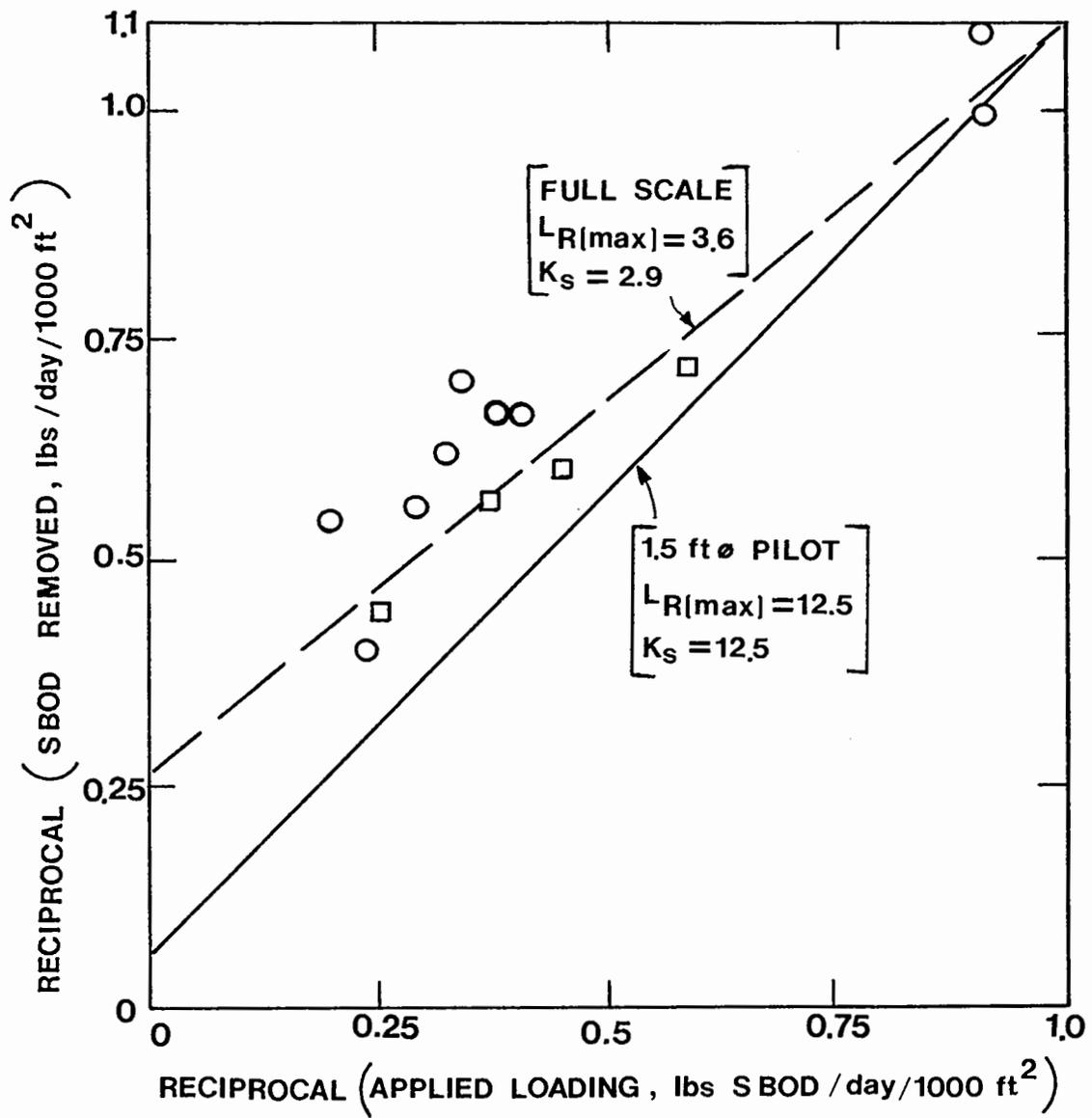
This equation can also be rearranged to make a linear plot, as follows:

$$\frac{1}{L_R} = \frac{K_S}{L_{R(\max)}} \cdot \frac{1}{L_0} + \frac{1}{L_{R(\max)}} \quad (2)$$

Now, $L_{R(\max)}$ and K_S can be determined from the slope and intercept.

When the reciprocals of the refinery and municipal data in Table I are plotted, the lines shown in Figure 7 are developed. The dashed line represents the full scale RBC data and the solid line represents the 1.5 foot RBC pilot data. The squares are the actual refinery data, and the circles are the actual municipal data. At an applied loading of 1.0 lb BOD/day/1000 ft² the BOD removed is seen to be the same for both the pilot and full scale systems. $L_{R(\max)}$ for the pilot and full scale systems are 12.5 and 3.6, respectively, while the respective K_S values are 12.5 and 2.9.

The full scale theoretical maximum BOD removal rate of 3.6 corresponds to the predicted loading point where actual zero order kinetics occurs. However, apparent zero order kinetics due to oxygen limited BOD removal becomes obvious



at loading rates around 4.0 to 6.0 lbs SBOD/day/1000 ft² corresponding to removal rates around 2.0 to 2.5 lbs SBOD/day/1000 ft², as can be observed in Figures 5 and 6. The Monod expression is based on limited BOD removal due to substrate saturation kinetics, not oxygen limitations, and therefore the reciprocal relationship shown in Figure 7 is only good up to loading rates around 6.0 lbs SBOD/day/1000 ft². Apparent zero order kinetics with a maximum removal rate of 2.5 lbs SBOD/day/1000 ft² are observed due to oxygen limitations, even though the Monod kinetics indicate the theoretical maximum removal rate of 3.6 lbs SBOD/day/1000 ft². Thus, this information from Figure 7 and equation one can be effectively employed to predict the amount of BOD removed for BOD loading rates up to about 6 lbs SBOD/day/1000 ft² for the full scale RBC's.

SCALE-UP AND OXYGEN TRANSFER CONSIDERATIONS

Figures 4, 5 and 6 show that treatment efficiency, in terms of lbs SBOD removed/day/1000 ft², of the RBC's investigated during this study was independent of media diameter below total loadings of about 1.0 to 1.5 lbs SBOD/day/1000 ft². BOD removals of 90 percent and greater were achieved at these and lower loading rates in all the RBC's. At soluble BOD loading rates greater than 1.5 to 2.0 lbs/day/1000 ft² the treatment efficiency of the full scale RBC's decreased faster than the pilot systems. Around 1.5 to 2.0 lbs SBOD/day/1000 ft² the full scale systems started becoming oxygen transfer limited instead of biochemical reaction rate limited with an observed maximum removal rate occurring at around 6 lbs SBOD/day/1000 ft² loading.

The full scale systems treating refinery and municipal wastewaters exhibited a theoretical maximum BOD removal rate of 3.6 lbs SBOD/day/1000 ft² of media with a substrate saturation constant of 2.9 lbs SBOD/day/1000 ft². However, the actual acceptable maximum removal rate of 2.5 lbs SBOD/day/1000 ft² occurred at an applied loading rate around 6.0 lbs SBOD/day/1000 ft². Thus, at this applied loading rate the BOD removal rate approached a constant value with increased loading, and the system was operating at apparent zero order reaction kinetics due to oxygen transfer limitations.

A major constraint in the design scale-up of an RBC

system then becomes the applied loading rate to the first stage(s) to values compatible with the system oxygen transfer capabilities. In the systems investigated in this study there were no oxygen limitations below 1.0 to 1.5 lbs SBOD/day/1000 ft² loading, indicating no problems in direct scale-up of full scale systems from pilot systems below these loading rates. However, at loading rates greater than 1.5 to 2.0 lbs SBOD/day/1000 ft² oxygen limitations become apparent and must be considered during scale-up from pilot studies. In order to avoid oxygen transfer problems during scale-up the first stage(s) must not be loaded over 1.0 to 1.5 lbs/day/1000 ft². Higher loadings will result in oxygen transfer problems and reduced removal rates. First stage(s) loadings greater than 2.0 lbs/day/1000 ft² will cause oxygen limited organics removal, and overall loadings of around 6.0 lbs/day/1000 ft² and greater will cause apparent zero order removal rates.

During evaluation of different diameter RBC's the limiting factor for substrate removal has been shown to be oxygen limitation. Smaller diameter RBC's provide better oxygen transfer and higher removal rates at higher organic loadings compared to larger diameter systems. Since oxygen limitation, not available surface area, is the limiting factor similar characteristics should exist in any fixed bed biological reactor. Unless the surface area available is not adequate to provide an adequate biomass to substrate ratio, the organic removal rate in any fixed bed reactor will be eventually oxygen limited instead of biochemical reaction rate limited. Kincannon (6) has shown similar analyses of plastic media biological towers to yield the same conclusions presented here for RBC's. When plotting BOD removed versus surface area or volume of filter media, the same types of plots were observed with the higher organic loadings yielding apparent zero order reaction kinetics.

In fact, plots of BOD removed in lbs/day/1000 ft² versus BOD applied in lbs/day/1000 ft² for RBC's and plastic media towers showed essentially the same removal characteristics. Thus, the organic removal rate characteristics per unit surface area for both these systems are similar. This analysis indicates that the RBC functions primarily as a fixed bed reactor with the primary mechanism of oxygen transfer being the rotation of the media through the air instead of direct oxygen transfer into the liquid. Oxygen transfer

from the gas phase into the liquid flowing over the media as the RBC rotates through the air is apparently the mechanism of oxygen supply for this biological wastewater treatment process. Heidman also suggests that the transfer of dissolved oxygen into the bulk liquid is minimal, and that the oxygen transfer into the biofilm during the air exposure cycle is by far the major contributor in satisfying the oxygen demand with a comparison of total oxygen transfer capability among units based on direct comparison of K_{1a} values to be potentially misleading (7).

As previously described, the oxygen transfer capabilities of a particular manufacturer's media have been standardized due to standardization of their systems physical requirements and constraints. The problem of scale up then simply becomes one of matching an RBC system oxygen transfer capabilities with the oxygen requirements of a particular wastewater. By knowing the oxygen limitations of the full scale system, the design engineer can determine the BOD removal versus BOD loading characteristics of that particular wastewater from pilot studies and match the desired treatment efficiency with the oxygen transfer capabilities of the full scale system. Therefore, the impetus should be on the equipment manufacturers to define and optimize the oxygen transfer capabilities or limitations of their systems. Transfer capabilities need to be defined in terms of specific media designs such as surface area requirements per unit volume of media since oxygen transfer into the biofilm during air exposure is the major contributor to satisfying the oxygen demand.

SUMMARY AND CONCLUSIONS

The total organic loading approach to design of RBC's previously developed by the authors in 1972, has been presented in detail and compared to other design approaches. The advantages and benefits of this RBC design method compared to other existing design approaches has been demonstrated. And yes, the question of which parameter, hydraulic loading, organic concentration or total organic loading, to use for proper design and operation of the RBC process has been answered.

Any design method based on zero, first, or second order

rate kinetics will be limited to loading conditions within a specific loading range for design of RBC's. Design in terms of first order reaction kinetics will require determination of two separate reaction rate constants. Depending on the operating condition or loading condition of an RBC, the kinetic removal rates through the system can be shown to follow apparent zero, first, or second order reaction rate kinetics. The total organic loading design method can accurately predict the amount of BOD removed (lbs BOD/day/1000 ft²) per applied BOD loading (lbs BOD/day/1000 ft²), irrespective of apparent zero, first, or second order reaction kinetics. The excellent fit of the Monod type relationship to BOD removed versus BOD applied allows prediction of BOD removal for any applied loadings desired by application of monomolecular kinetic analysis.

Monod kinetics also predicts the maximum BOD removal rate observed in RBC's. At a certain loading condition, the RBC becomes saturated with BOD, apparently due to oxygen limitations, and the removal rate does not increase with increasing BOD loadings. The system becomes oxygen limited at these loading conditions and exhibits apparent zero order kinetics. As the RBC diameter increases, the rotational speed and oxygen transfer capabilities decreases. Smaller diameter systems transfer more oxygen, and thus their maximum BOD removal rates are greater than larger systems, as the previous data shows.

The primary mechanism of oxygen transfer in RBC systems appears to be transfer from the gas phase into the liquid flowing over the media during rotation. The submerged portion of the media does not accomplish effective oxygen transfer but does provide contact retention time for the wastewater and equalization capacity. The tradeoffs of maximum exposed surface area for maximum oxygen transfer, surface area per unit volume of media, and liquid volume are key variables related to mass transfer of oxygen in the RBC process. To effectively accomplish scale-up of RBC's from pilot data, the oxygen transfer capabilities and limitations of the full scale systems must be defined.

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PART XV: EXPERIENCES WITH FIXED FILM TREATMENT
FACILITIES

RBC SUPPLEMENTAL AIR:
CONTINUOUS OR INTERMITTENT?
YOUGHIOGHENY WASTEWATER TREATMENT PLANT
NORTH HUNTINGDON TOWNSHIP, PENNSYLVANIA

JEFFREY W. HARTUNG. (SYSTEM MANAGER)

The purpose of this report is to discuss the results of continuing research being conducted on the effects of supplemental aeration by comparing continuous application to intermittent application with regards to bio-disc loading and the protection of the bio-disc equipment.

INTRODUCTION

In 1969, North Huntingdon Township Municipal Authority completed the construction of a 1.5 million gallon per day (mgd) intermediate treatment plant (halfway between primary treatment and secondary treatment - 50% removal BOD₅ and 50% removal suspended solids). During the construction of the plant, the Sanitary Water Board of Pennsylvania amended its regulations to thereafter require minimum treatment for all discharges of bio-degradable waste to be secondary treatment, except for some highly acid receiving streams. Shortly after the start of the plant operations, the Sanitary Water Board formally notified the North Huntingdon Township Municipal Authority of the required performance modifications to its original permit and ordered that it proceed expeditiously to

plant is shown on Plate #1.

HISTORY

During the six years the bio-discs have been in operation, there has been six major maintenance problems, all of which had occurred in the first five years before the supplemental air was added. The problems were as follows:

1. February, 1977 - Cracked shaft on #1 Bio-disc.
2. April, 1978 - Cracked shaft on #2 Bio-disc.
3. May, 1978 - End bearing movement on Bio-disc #4.
4. May to September, 1978 - All shafts were replaced and bearing on #4 was replaced.
5. July, 1980 - Bearings on Bio-disc #2 and Bio-disc #4 moved.
6. October, 1980 - Media collapsed on Bio-disc #1.

The specific reasons for each breakdown cannot be fully determined, but in each case, one of the underlying factors may have been an excessive weight problem. In discussing the problem, we concurred with the manufacturer and other plant personnel with similar set-ups and problems. It was established that an overloaded bio-disc, that which achieves weight heavier than manufacturers suggested total weight, may encounter the aforementioned problems.

This overload condition may occur when the loading to the Bio-disc is too great (BOD-TSS) causing excessive growth of bio-mass and from trapped inorganic materials that become lodged in the media itself. During overloaded conditions, the overall performance of the bio-disc units are poor. Microscopic examination of the bio-mass shown anerobic growth (beggiatoa) forming heavily between the bio-mass and the plastic media, and a great reduction in feeders (rotifers and ciliates). October, 1980, supplemental air was chosen as a remedy to prevent excessive bio-mass growth from occurring. By diffusing air beneath the bio-disc units, turbulence will be created to assist the sheering forces of the bio-mass. With the assistance of our Consulting Engineers; Betz, Converse and Murdoch, we were able to acquire two Roots, positive displacement blowers from a dismantled plant of a nearby municipality. The blowers are type AF, size 65, powered by a 7.5 hp; 1760 rpm motor, with a 1.714 ratio from motor to blower. Capable of producing 185 cubic feet per minute (cfm) at 2 pounds per square inch (psi) each. A diffuser header was fabricated out of 2" diameter Schedule

40 PVC pipe with two rows of 30 diffusing holes $\frac{1}{2}$ " diameter, spaced $8\frac{1}{4}$ ". The diffuser header was then anchored to a 3" steel channel and mounted directly under the center line of the bio-disc with the diffusing direction slightly angled against rotation of the bio-disc. Shut off valves and pressure gauges were placed within the entire system at predetermined locations to offer flexibility throughout the system to adjust the volume and pressure of the air to any one or all of the bio-discs.

In December, 1980, the entire supplemental air units were in place and put into operation at a total cost of material and labor at \$6,000.

The initial mode of operation was to diffuse a constant volume of air continuously to all the units. During the continuous phase of operation, air is diffused under the two end units (1 and 4) with 110 cfm at 2.5 psi while the middle units (2 and 3) received 125 cfm at 1.75 psi. Due to a suspected O_2 depletion of the first unit, the second and third units will require a greater amount of diffused air to sustain a healthy culture. Within four to five weeks a healthy culture was established and again examined periodically with the microscope of which we continually found a healthy population of feeders, rotifers and ciliates and a zero population of beggiatoa. It was during these early observations that we noticed the excited activities of the younger feeders. Naturally, it's known that the healthier the bio-mass the more effective it will be. After several months passed of the continuous phase of operation, we questioned our application of continuous diffused air with the possibility of discontinuing the continuous mode and initiating an intermittent mode of diffusing large volumes of air to blast off all the matured bio-mass growth to allow a new culture of the young and more active feeders to grow. During this intermittent mode, a large volume of air (370 cubic feet per minute (cfm) at 2 pounds per square inch (psi) is diffused to each individual bio-disc for thirty minutes, starting at unit #1 and proceeding to unit #4 (therefore allowing the gravity flow of the sewage to carry the bio-mass to the effluent), scouring off all organic and inorganic matter. Within 4-5 weeks, an entirely new culture will be matured and the preceding method of scouring will again be administered.

May of 1981, we began to administer the intermittent mode of diffused air. Microscopic examination of the biological activity confirmed our beliefs that intermittent replacement of the bio-mass enhances a healthy, active bio-

logical growth. An important factor we also considered is operating cost. Measuring the amperage draw of our motors, we have found each motor drawing 6 amps. To determine the kilowatt per hour (kwh) we used the following formula:

$$(4.56) \text{ kw} = \frac{1.73 \times 6 \text{ (amps)} \times 440 \text{ (volts)} \times 1 \text{ (power factor)}}{1000}$$

Therefore, each blower consumes 4.56 KWH. Our supplier (West Penn Power) charges .0375 dollars per KWH. Plugging these factors into a formula for yearly cost we find:

$$(4.56) \text{ kwh} \times 2 \text{ (blowers)} \times 8760 \text{ (hours per year)} \times \$0.0375 = \$2995.92 \text{ per year for continuous operation.}$$

As described in the intermittent mode of operation, the blowers are run for only 2 hours per month (30 minutes each times four discs). So by the same formula for a yearly cost we find:

$$4.56 \text{ kwh} \times 2 \text{ (blowers)} \times 24 \text{ (hours per year)} \times \$0.0375 = \$8.21 \text{ per year for intermittent operation.}$$

CONCLUSION

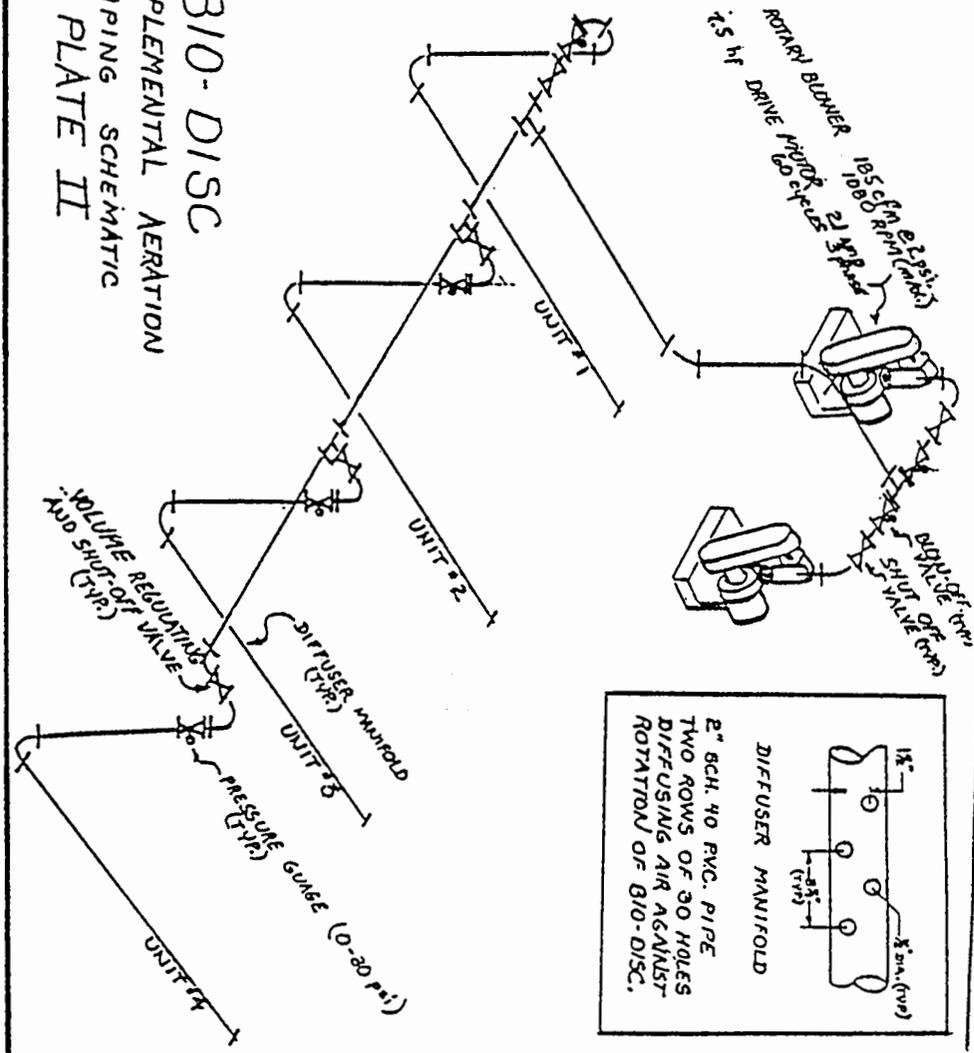
Intermittent replacement of the entire biological growth on the bio-disc media, by the application of diffused air, enhances the growth of young more active bio-mass feeders; controls the excessive build-up of bio-mass; prevents anaerobic growth from occurring; maintains a total weight of the equipment in a safe limit to prevent mechanical failures; and prolongs the life of the equipment, at an inexpensive operational cost.

Before implementation of this technique, consideration must be given to the secondary clarifier to assure the retention time of the bio-disc effluent is sufficient to allow proper settlement of the shock load of bio-mass material. In our particular case, there was sufficient settlement time.

As mentioned in the beginning of this report, these techniques of supplemental air are part of continuing research. We are still in the process to determine the BOD₅ and TSS removal efficiency of various applications of supplemental air; however, we have established the tentative

finding that continuous application or intermittent application has not noticeably affected our effluent quality that continues to remain well within our N.P.D.E.S. parameters of 30 mg/l BOD₅ and 30 mg/l TSS throughout the year.

BIO-DISC
 SUPPLEMENTAL AERATION
 PIPING SCHEMATIC
 PLATE II



THE OPERATOR'S VIEWPOINT OF WASTEWATER TREATMENT
USING ROTATING BIOLOGICAL CONTACTORS

Mary A. Bergs, P.E. Davy Engineering Company,
La Crosse, Wisconsin.

INTRODUCTION

Rotating biological contactors, also known as RBC's, RBS, RBD, rotating bio-discs, bio-shafts, etc., are a relatively new wastewater treatment process. Laboratory studies have been performed to study the kinetics of the process. Manufacturers each claim their own design is the best. Engineers have selected RBC's for some treatment facilities because of low operation and maintenance costs. Generally, laboratory studies have been done on bench scale models and the researchers can stop the experiment when they wish. Even though they may follow-up after a sale or construction, manufacturers and engineers generally are not involved with the daily operation of RBC's.

The purpose of this report is to present the viewpoint of the wastewater treatment plant operator, the person who must operate the RBC treatment facility efficiently 24 hours a day, 7 days a week.

DESCRIPTION OF FACILITIES

Operators representing one-third of the municipal wastewater treatment facilities using RBC's in Wisconsin were interviewed. No two of the facilities are exactly alike. A description of each facility follows.

Augusta, Wisconsin

One four-stage mechanical drive RBC shaft was used to upgrade a trickling filter facility in Augusta, Wisconsin. The existing facility consisted of a primary clarifier, 37 foot diameter trickling filter, final clarifier and anaerobic digester. The former final clarifier is now used as an intermediate clarifier prior to the RBC, new final clarifier, chlorination and post aeration facilities. During upgrading, the rock filter media was replaced with plastic media. The trickling filter, RBC, and final clarifier each have a cover. The facility is designed so that some or all of the primary clarifier effluent can be bypassed directly to the RBC.

The RBC has 114,400 square feet of media. Baffles separate the four sections of media. The last two sections are high density media. Start-up of the RBC was April 9, 1980.

The facility is designed for a flow of 333,600 gallons per day (GPD) including 91,600 GPD of infiltration. Actual flows have ranged from 100,000 GPD in winter to an average of 150,000 GPD during the rest of the year. No industries discharge wastewater to the treatment facility. Effluent requirements for the facility are 20 milligrams per liter (mg/l) BOD₅, 20 mg/l suspended solids, 16 mg/l ammonia-nitrogen and 7 mg/l dissolved oxygen in the winter and 30 mg/l BOD₅, 30 mg/l suspended solids, 32 mg/l ammonia-nitrogen and 7 mg/l dissolved oxygen in the summer.

Delafield-Hartland, Wisconsin

The treatment units at the Delafield-Hartland wastewater treatment facility are an aerated grit chamber, two primary clarifiers, 28 mechanical drive RBC shafts, two final clarifiers, shallow bed sand filters, chlorination, cascade

aeration at the outfall and one primary and one secondary anaerobic digester. One primary clarifier, 14 RBC shafts, and one final clarifier are not presently in use.

The RBC shafts are arranged in four rows of 7 shafts each for a total of 850,000 square feet of media in each row. Baffles separate all but the first two shafts in each row. The last three shafts in each row are high density media. The RBC's are housed in a block and brick building.

The facility is designed to treat 2.2 million gallons per day (MGD). Since its July 29, 1980, start-up, the facility has been treating 0.8 to 1.2 MGD of domestic wastewater and 1,000 GPD of leachate from a nearby sanitary landfill. Effluent limits are 15 mg/l BOD₅, 15 mg/l suspended solids and 4.5 mg/l ammonia-nitrogen in the winter and 10 mg/l BOD₅, 10 mg/l suspended solids and 2 mg/l ammonia-nitrogen in the summer.

Eau Claire, Wisconsin

Fifty-six air-drive RBC shafts were used to upgrade a primary facility at Eau Claire. Other treatment units at the facility are an aerated grit chamber, four primary clarifiers of which three are presently being used, three final clarifiers with two in use, chlorination, a sludge thickener and two primary and two secondary anaerobic digesters. Forty of the RBC shafts are presently in use.

Flow to the RBC's is along a central channel. Seven rows of 4 shafts each are along either side of the channel with baffles separating the 4 shafts in each row. Each shaft has 100,000 square feet of media. Each shaft has a fiberglass cover.

The facility is presently receiving 5.2 MGD of combined domestic and industrial flow. Design flow is 16.3 MGD. Effluent limits are 30 mg/l BOD₅, and 30 mg/l suspended solids.

Fennimore, Wisconsin

The wastewater treatment facility at Fennimore consists of a stormwater holding pond, aerated equalization tank, two primary clarifiers, two RBC "units", two final clarifiers,

anthracite filters, chlorination and an anaerobic digester. Because of the arrangement of the duplicate units, one primary clarifier, RBC unit and final clarifier can be operated independently of the matching set of treatment units. The equalization tank and final clarifiers are each covered with an aluminum dome.

Each RBC "unit" consists of 4 air-drive RBC shafts separated by baffles. The fourth shaft in each unit is high density media. There are 450,000 square feet of media in each RBC unit. Each of the RBC shafts has a fiberglass cover.

The treatment facility receives flow from a dairy, slaughtering operation and other industries in the City. Since start-up of the RBC's on January 25, 1980, the treatment facility has treated flows that have been slightly higher than half of the design flow of 620,000 GPD. Less than a year after start-up, the facility was treating the organic loading that the facility was designed to treat. After the City began detailed surveillance monitoring of industrial discharges and strict enforcement of the sewer use ordinance, the organic load dropped back to approximately half of the design load.

The facility is required to meet effluent limits of 15 mg/l BOD₅, 20 mg/l suspended solids and ammonia-nitrogen of 3 mg/l in summer and 6 mg/l in winter.

Fountain City, Wisconsin

Two RBC shafts, a final clarifier and chlorination facilities were added to two primary clarifiers and an anaerobic digester to upgrade the Fountain City wastewater treatment facility. The RBC shafts and the final clarifier have individual covers.

Each of the mechanical drive RBC shafts has two stages separated by baffles. The 20 foot shafts can be operated either in series or in parallel. The RBC's have been operated in parallel since the April 6, 1981 start-up. The shafts have a total of 152,000 square feet of media.

The facility treats an average of 112,000 gallons per day which is about half of the design flow. No industrial flow is treated at the facility. Effluent limits for the facility are 30 mg/l BOD₅ and 30 mg/l suspended solids.

Lancaster, Wisconsin

Eight air-drive shafts were used to upgrade the wastewater treatment facility at Lancaster. Treatment units at the facility include an aerated stormwater holding pond, aerated grit chamber, new primary clarifier, 100 foot diameter rock media trickling filter, the RBC shafts previously described, two final clarifiers, chlorination and an anaerobic digester. The original primary clarifier can be used as a standby unit. Upgrading included putting a cover over the trickling filter.

The eight RBC shafts are arranged in two rows of four shafts with baffles between each shaft. To prevent short-circuiting, the first and second and the third and fourth shafts are rotated towards each other. The two rows are operated in series. The fourth shaft in each row is high density media; the total media area is 920,000 square feet. A frame building of treated wood covers the RBC shafts. Start-up of the RBC shafts was in April of 1979.

The facility is designed to treat 740,000 gallons per day to meet effluent limits of 15 mg/l BOD₅, 20 mg/l suspended solids and ammonia-nitrogen of 3 mg/l in the summer and 6 mg/l in the winter. It is presently treating 500,000 gallons per day including wastewater from a cheese factory in the City.

Union Grove, Wisconsin and Eagle Lake, Wisconsin

The Union Grove wastewater treatment facility has a contact stabilization plant followed by three mechanical drive RBC shafts, two pressure sand filters and chlorination.

The three RBC shafts are operated in series with baffles between the shafts. Each shaft has 160,000 square feet of media. The RBC's were first started January 29, 1979. Mechanical problems, which are described later in this report, occurred after start-up. This is the first time that the RBC's have been operated through a winter.

Designed for a flow of one MGD, the facility is presently receiving 550,000 gallons per day of domestic wastewater. Effluent limits are 15 mg/l BOD₅, 20 mg/l suspended solids, 1 mg/l phosphorus and ammonia-nitrogen of 6 mg/l in the winter and 4 mg/l in the summer.

The operator of the Union Grove facility is also the operator at the Eagle Lake wastewater treatment facility. The Eagle Lake facility is very similar to the Union Grove facility except that it is smaller and does not have tertiary filters. The Eagle Lake facility began operating about six months before the Union Grove facility.

OPERATOR COMMENTS

The amount of operation and maintenance required, and the ease with which it is performed, depends on factors such as the manufacturer's design of the RBC, type of enclosure, and the type of treatment the RBC is designed to provide. The various factors are discussed separately below, but they are inter-related.

Manufacturer

Augusta, Eau Claire, Fennimore, Fountain City and Lancaster all have RBC units manufactured by Autotrol. The RBC's at Eagle Lake and Union Grove were manufactured by Bio-Shaft. Delafield-Hartland has RBC's manufactured by Envirodisc.

Requirements for lubrication of bearings, motors and blowers are similar for the three manufacturers. The operators at Fennimore and Lancaster lubricate the shaft bearings twice a week. At Union Grove, Eagle Lake, Augusta and Fountain City the shaft bearings are lubricated once a week. The shaft bearings are lubricated once every two weeks at Eau Claire and once every three weeks at Delafield-Hartland. Other motor and blower lubrication is generally done yearly.

The RBC's at Delafield-Hartland are made up of pie-shaped sections of media. The media is carried within a metal framework. Either because of wear or compaction, after a period of time the media is slightly smaller than the framework. Braces holding the media can be tightened to keep the media from slipping around inside the framework. There are two bolts on each end of the two braces holding each pie-shaped section of media. It took 56 manhours to tighten the braces on one shaft. Because of experience and better equipment, only 20 manhours were needed to tighten the braces on

other shafts. The manufacturer has provided bolt holes to tighten the braces twice during the life of the discs. The only other mechanical problem with these RBC's so far is that several motor drive belts have broken and had to be replaced.

Operation of the Autotrol units is slightly different for the mechanical drive and air-drive units. Operators at Augusta and Fountain City reported no mechanical problems to date. It is known that there have been shaft failures at some other installations. The operator at Eau Claire did not report any mechanical problems. The Fennimore facility originally had variable speed blowers. Because the belts burned up very quickly, the blowers were replaced with constant speed units. The operator at Lancaster reported that drive belts have cracked on the blower which is run at a higher rpm. A notched belt has recently been installed to see if this will solve the problem. The operators did not report any problems with diffusers clogging. However, the operator at Fennimore said it would be difficult to tell if only one or two diffusers were clogged because of the turbulence of the rotating shafts.

The Union Grove facility has had several problems. A gear drive failure occurred three weeks after start-up because of erroneous instructions from the manufacturer. All of the tie rods broke, were fixed and then broke again. They were then fixed again. During construction, the RBC's were set in the tank crooked. This was a contributing factor to problems of sprocket and chain alignment. Soft gears wore down and had to be replaced. Since the Eagle Lake facility is so similar to the one at Union Grove, any repairs made at one facility were also made at the other even if the problem was not apparent there. The RBC's have not had any mechanical problems the past few months.

Type of Enclosure

Every operator mentioned the RBC enclosure even though it was not a standard interview question. Even those operators who had only good things to say about RBC's mentioned something they did not like about the housing for the RBC's. The two types of enclosures for the RBC's described in this report were a building housing all the RBC shafts or individual fiberglass covers over each shaft.

Some of the disadvantages of the individual covers would be reduced in a warmer climate. In winter, the snow must be shoveled away from the doors to each shaft. Because of the warm humid air inside the cover, the doors have a tendency to freeze shut and the walk-way in front of the door becomes coated with thick ice. There is usually not enough room for a man to be inside the enclosure when performing maintenance. The operators do not like lubricating bearings while standing outside in below zero weather. The problems are multiplied by the number of shafts to be serviced. Even if a man can fit inside the enclosure, the doors are only 3 to 4 feet high. The operators would prefer 6 foot high doors. Portholes on the individual covers are too small to get a good look at what is happening along the length of the shaft. The required tightening of braces at the Delafield-Hartland facility would be impossible with individual covers on the shafts.

The disadvantages of individual enclosures are solved when RBC's are put inside a building. But the buildings also have disadvantages. The roof of the pole building at Lancaster is designed so that a section can be removed if it is ever necessary to remove or replace a shaft. Most buildings make it difficult or impossible to replace a shaft if it is necessary. The buildings are very humid inside. The humidity has caused problems such as filter flies and spiders at the Union Grove facility, moisture shorting out electrical circuits and mold growing on the concrete block walls at the Delafield-Hartland facility, and water dripping, or raining, from the ceiling causing slippery floors in each RBC building. The dripping water has not affected growth on the media. Heating and/or cooling and ventilation of the buildings is difficult and expensive. Splash guards were added to the Delafield-Hartland facility to prevent the RBC's from splashing water on the floor and causing slippery floors.

Type of Treatment

The RBC's at Fountain City and Eau Claire are only designed for BOD removal. Because of the low loadings at both facilities, some nitrification is occurring. The RBC's provide only nitrification at the Union Grove and Eagle Lake facilities. All of the others are designed to provide BOD removal and nitrification. All of the facilities are presently meeting effluent limits.

The operator at Union Grove has tested the RBC influent and effluent. The BOD and suspended solids are approximately the same in the influent and effluent. The ammonia-nitrogen is significantly less and the dissolved oxygen significantly higher in the effluent.

The Lancaster, Fennimore and Eau Claire facilities treat both domestic and industrial flow. The three facilities have had different experiences with industrial flows. The operator at Fennimore said the influent has been white from milk and a few minutes later red with blood from the slaughtering operation and there was no noticeable effect on the RBC's. Effluent quality remained good. An accidental spill of cleaning fluid from the dairy stripped the growth from the RBC's in Lancaster. The spill occurred at the end of March 1981 and it took several months for the plant to return to normal. Because of the dairy, the pH of the wastewater fluctuates considerably. Calcium deposits are developing on the RBC's. Industries in Eau Claire notify the treatment facility when a spill occurs. Because they are concerned about killing the growth, the RBC's are bypassed when an industrial spill occurs.

Start-up

The operators were asked "How long after start-up did it take to get noticeable growth on the RBC media?" and "How long after start-up did it take until the RBC units provided treatment sufficient to meet effluent limits?" As expected, the time of year that start-up occurred had a definite effect on the answers. Start-up in January took about twice as long as start-up in April or October.

The Delafield-Hartland facility received about 300,000 gallons per day of effluent with a BOD₅ of 50 to 60 mg/l from the old Hartland treatment facility for several weeks before start-up. Growth could be felt but not seen on the black media. The RBC's were then shut off for several days to allow completion of construction. It took one week after start-up with raw wastewater before the facility was meeting effluent limits. Start-up was in July.

Starting about Thanksgiving, the Fennimore wastewater treatment facility received 80,000 GPD of raw wastewater. One row of RBC's was used. Start-up with an additional 260,000 GPD occurred the end of January. The facility did not meet effluent limits until the end of the following March.

Operation

There is little the operator can do in-plant to increase the efficiency of the RBC. The rotational speed of the RBC's, return of digester supernatant and sludge withdrawal from the final clarifier are the three operational items that affect effluent quality. Alternate methods of operating the treatment facility, such as series and parallel operation of the RBC units, is not possible at all treatment facilities.

The first stage RBC's at Eau Claire, Lancaster and Fennimore are rotated faster to encourage sloughing and avoid overloading the shafts. First stage speeds range from 1.2 to 1.6 revolutions per minute (rpm). The speed of the last stage is 1.0 to 1.1 rpm. The faster speed of the first stage also seems to prevent an uneven rotation of the RBC, called loping by the Eau Claire operator.

Loping has occurred at all three air-drive facilities, but at none of the other facilities. The operator at Eau Claire kept track of how long the loping lasted. They tried increasing the rotational speed, decreasing the rotational speed, shutting off flow to the shaft to starve it and not doing anything. In all cases, it took about a week for the loping to go away. The two lead shafts at Fennimore were out of balance during the 1980-81 winter until the raw wastewater temperature increased to 50°F. Part of the problem may have been high organic loadings from industries.

The operator at Fennimore said that it only takes half as much air to drive the shafts in summer, when the raw wastewater temperature is above 60°F, than in winter, when the raw wastewater temperature is 45°F. This was not noticed at the other two air-drive facilities.

The facilities with mechanical drive RBC's operate them at 1.3 to 1.6 rpm. All the shafts at a given facility are rotated at the same speed. At Delafield-Hartland, one set of RBC's was rotated at 1.2 rpm and one set at 1.6 rpm to see what would happen. The only noticeable difference was the effluent alkalinity. At 1.2 rpm, the alkalinity was 305 mg/l; the alkalinity was 315 mg/l at 1.6 rpm. Influent alkalinity is 400 mg/l.

Periodic return of digester supernatant does not affect growth on the RBC's. The operators do notice an increase in effluent BOD and suspended solids when supernatant is returned. Sludge from the digesters at Eau Claire was returned continuously while the digesters were not working. During

that time the growth on the RBC's was thicker and black in color. One of the 4 baffle boards between the first and second shaft in each row of RBC's was removed to avoid overloading the first stage. The boards were not replaced.

At Fountain City and Augusta, sludge is removed from the final clarifier once a day. The sludge was thinner when it was removed more often. At Lancaster, Delafield-Hartland, and Eau Claire, sludge is removed from final clarifiers continuously or at least once per hour on an automatic pumping cycle. Twice a day removal was tried at Lancaster but denitrification occurred in the clarifier. Once a day removal was tried at Delafield-Hartland, but too much sludge accumulated. The sludge was also thicker. The operator at Fennimore sets telescopic valves to remove sludge at the rate of about 40 gallons per minute from the final clarifier. Because sludge seemed to build up in the clarifiers, twice a day sludge is removed at about 300 gallons per minute for 15 to 20 minutes. The continuous removal limits denitrification, but a thicker sludge is obtained with short periods of high rate sludge removal. Union Grove and Eagle Lake do not have clarifiers following the RBC's.

The operator does not have control over the wastewater temperature, but it appears to affect operation of the RBC's. The operator at Fennimore has kept detailed records of the air temperature, wastewater temperature and the effect on the RBC's. He reported that air temperature does not seem to have much effect on the RBC's. Sloughing increases with a 1 to 2^oF change in water temperature. The RBC's were almost cleared of growth in the spring. However, spring flows are also higher due to infiltration and inflow. The RBC's have their heaviest growth in the winter. The operators at Union Grove and Augusta reported a heavier growth on the RBC's this winter starting about Thanksgiving. Augusta has lower flows in the winter when infiltration and inflow are at a minimum.

Facility Design

At some facilities, a nuisance or inadequate treatment was caused by the facility design.

Filter flies are a problem inside the RBC building at Union Grove. Paper covered insulation is attached to the inside walls. Because of the insulation, the walls cannot be sprayed with chlorine to get rid of the filter flies.

The RBC's in Eau Claire receive uneven loadings because of the hydraulics of the feed channel. A "back pressure" develops in the channel. The center rows of RBC's receive the most flow and therefore the highest loading. At first, only the RBC's on one side of the channel were in use. Extreme loping occurred on the lead shaft in the center row of RBC's. Slight loping was noticed on the lead shaft of the end rows and as far back as the third shaft in the center row. The problem has been alleviated for now by using shafts on both sides of the feed channel. The two rows of RBC's at the near end are not presently in use. The problem may reoccur when the facility reaches design capacity.

The operator at Eau Claire would also like to have the flexibility of operating at least some of the rows in series. Partially treated wastewater could then be returned for more treatment while starting up a new row of RBC's.

At Lancaster, the baffle boards were not tied together at first. The motion from the rotating shafts caused the boards to bow out. Flow was short-circuiting between the baffle boards.

Other Comments And Observations

The trickling filter at Lancaster has green growth in spring and summer. In winter the growth is pink to white. The plastic media in the trickling filter at Augusta appears to have a growth similar to that on a first stage RBC shaft.

Growth on the RBC's at the various facilities was as described in textbooks and literature. It was thicker and a dark brown to grey color on the first shafts. On subsequent shafts, the growth was thinner and lighter in color. On the final stages, growth rarely covered the entire media and was golden brown in color. The shafts at Union Grove all had growth that was very thin and light brown. It was easier to see the growth on white media than on black media.

Some operators looked at the biomass under the microscope. At Eau Claire, stalked ciliates were observed in the biomass from the first and second stage shafts. Worms were seen in the biomass from the second, third and fourth stage shafts. At Delafield-Hartland, stalked ciliates predominated in the first stage biomass. Rotifers, ciliates and worms were found on later stages. At Union Grove, rotifers were thriving on all the shafts.

CONCLUSIONS

Parts of this report may sound quite negative, but the operators rated the performance of RBC's as good to excellent. The operator at Lancaster added the provision "until problems occur." Maintenance was rated as simple to average.

The facilities described in this report have been meeting effluent limits, which is the first criteria in evaluating a wastewater treatment facility. As many of the facilities are underloaded, only time will tell if they will continue to meet effluent limits.

RBC's have been promoted for their low operation and maintenance requirements. Some aspects of RBC design nullify this advantage. Lubrication requires a considerable amount of operator time in the larger facilities. Tightening the braces on the RBC's at Delafield-Hartland requires a major investment in operator time.

Quality control during manufacture of the RBC's and during installation would have reduced the problems experienced by the Union Grove facility.

Enclosures for RBC's are sometimes an afterthought in the design process. Access to the RBC's can have a significant effect on how often and well an operator performs necessary RBC maintenance. An operator is not likely to spend extra time checking the RBC's when he must stand out in below zero weather to do so. The cost for operation of a facility will be affected by any costs for heating, cooling and ventilating RBC enclosures.

The operator has very few options for running an RBC facility. Some of these options are eliminated during the design process. It is important to consider flexibility of operation during design.

Loping appears to occur when air-drive shafts are organically overloaded. A faster rotational speed encourages sloughing and helps to prevent loping. Once loping occurs, elimination of the overload conditions is the best solution.

The temperature of the raw wastewater has a significant effect on the RBC process. A small change in the temperature of the wastewater will induce sloughing of the biomass. Start-up takes considerably longer in cold weather.

Variation in the rotational speed of the RBC's does not appear to have a significant effect on the effluent. A faster rotational speed will promote sloughing which is important to prevent overloading of the first stage shafts. A slower

rotational speed may be suitable for later stages and will save energy.

RECOMMENDATIONS

Manufacturers should make sure their design does have the low maintenance requirements RBC's are reported to have. If tightening of braces is necessary, a faster and safer method than required on the units at Delafield-Hartland should be devised. As with any product, quality control is needed during manufacture and installation of RBC's.

A better enclosure is needed. Larger viewing ports along the side and room on the ends for the operator to walk inside when performing maintenance would be improvements to the present fiberglass covers.

Flexibility of operation should be stressed during design. The design should include methods for easily changing the rotational speed of the RBC's, flexibility in the frequency and amount of sludge removed from the final clarifiers and possibly a holding basin for digester supernatant so that it can be returned to the treatment process over a longer period of time. Consideration should be given to allowing both series and parallel operation of RBC's where more than one shaft is used.

Having twice as many shafts in the first stage as in the second stage would help prevent overloading of the first stage shafts.

Where possible, start-up should be planned for early summer.

INTERVIEWS

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Herrick, Jerry. Augusta Wastewater Treatment Facility,
Augusta, Wisconsin. Interview, 8 February, 1982.

Hyde, Robert. Delafield-Hartland Wastewater Treatment Facility,
Delafield, Wisconsin. Interview, 12 February, 1982.

Piel, Armin. Fountain City Wastewater Treatment Facility.
Fountain City, Wisconsin. Interview, 20 January, 1982.

Rosemeyer, Tony. Fennimore Wastewater Treatment Facility,
Fennimore, Wisconsin. Interview, 4 February, 1982.

TROUBLESHOOTING AN EXISTING RBC FACILITY

B. W. Newbry, Stanley Consultants, Inc.,
Muscatine, Iowa

M. N. Macaulay, Stanley Consultants, Inc.,
Muscatine, Iowa

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INTRODUCTION

Kirksville, Missouri, located in the north central part of the state, operates a rotating biological contactor (RBC) facility for municipal wastewater treatment. This plant serves a population equivalent of approximately 27,000, in terms of organic loading, and was the first facility of its type in the state of Missouri.

Shortly after going on-line in August, 1976, operation and maintenance problems were experienced. These included periodic effluent biochemical oxygen demand (BOD) concentrations in excess of discharge permit limitations, and structural failure of several RBC media central support shafts. In spite of efforts to remedy the situation, these problems continued, and increased in severity.

Upon the recommendation of the Missouri Department of Natural Resources (MDNR), the city retained a consultant to perform a detailed analysis of the plant's design, operational problems, and performance. The purpose of this study was to develop a plan for upgrading the plant to meet its

wastewater treatment objectives. This paper presents the results of a portion of that study.

The Kirksville Wastewater Treatment Plant (WWTP) uses the RBC modification of the fixed-film biological process for its municipal wastewater treatment. The plant includes primary settling, RBCs, and secondary settling. Secondary sludge is recirculated to the influent wet well and wasted with primary sludge to anaerobic digesters. A process schematic is shown in Figure 1. The treatment facility flow rate, BOD and SS design criteria, and discharge limits are given in Table I.

The biological treatment process of the Kirksville WWTP is shown in Figure 2. The system consists of 20 RBC units arranged in four parallel bays of five units in series. All units are mechanically driven without supplemental aeration. The nominal media surface area for each unit is 96,000 square feet. The shaft rotational speed is approximately 1.6 rpm. All RBCs are housed in a single control building.

Table I
Kirksville WWTP Design and Performance Data

Design Criteria

Flow

Average Day:	5.0 MGD
Maximum Day:	12.3 MGD

BOD₅

Raw Wastewater:	300 mg/l
Primary Effluent:	200 mg/l
Final Effluent:	20 mg/l

Suspended Solids

Raw Wastewater:	250 mg/l
Primary Effluent:	130 mg/l
Final Effluent:	20 mg/l

Effluent Limits

BOD ₅ (Daily Average):	30 mg/l
Suspended Solids (Daily Average):	30 mg/l

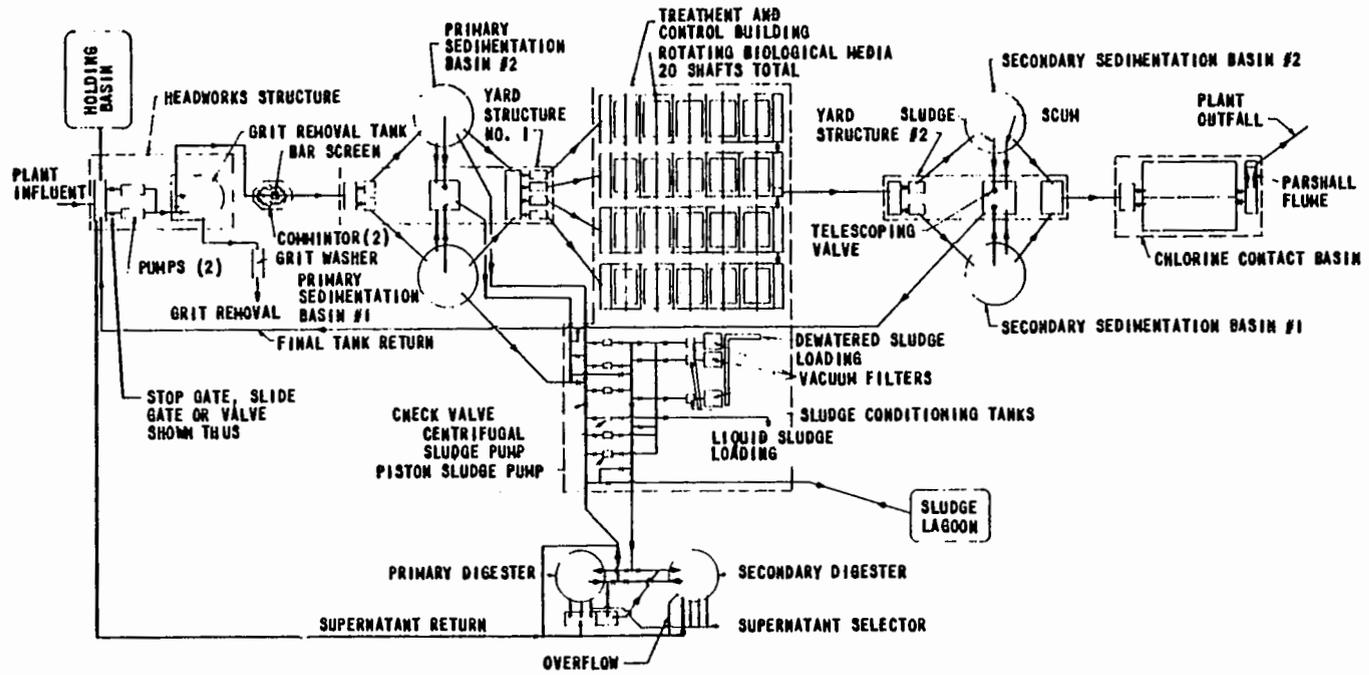


Figure 1. Kirksville WWTP Schematic Flow Diagram

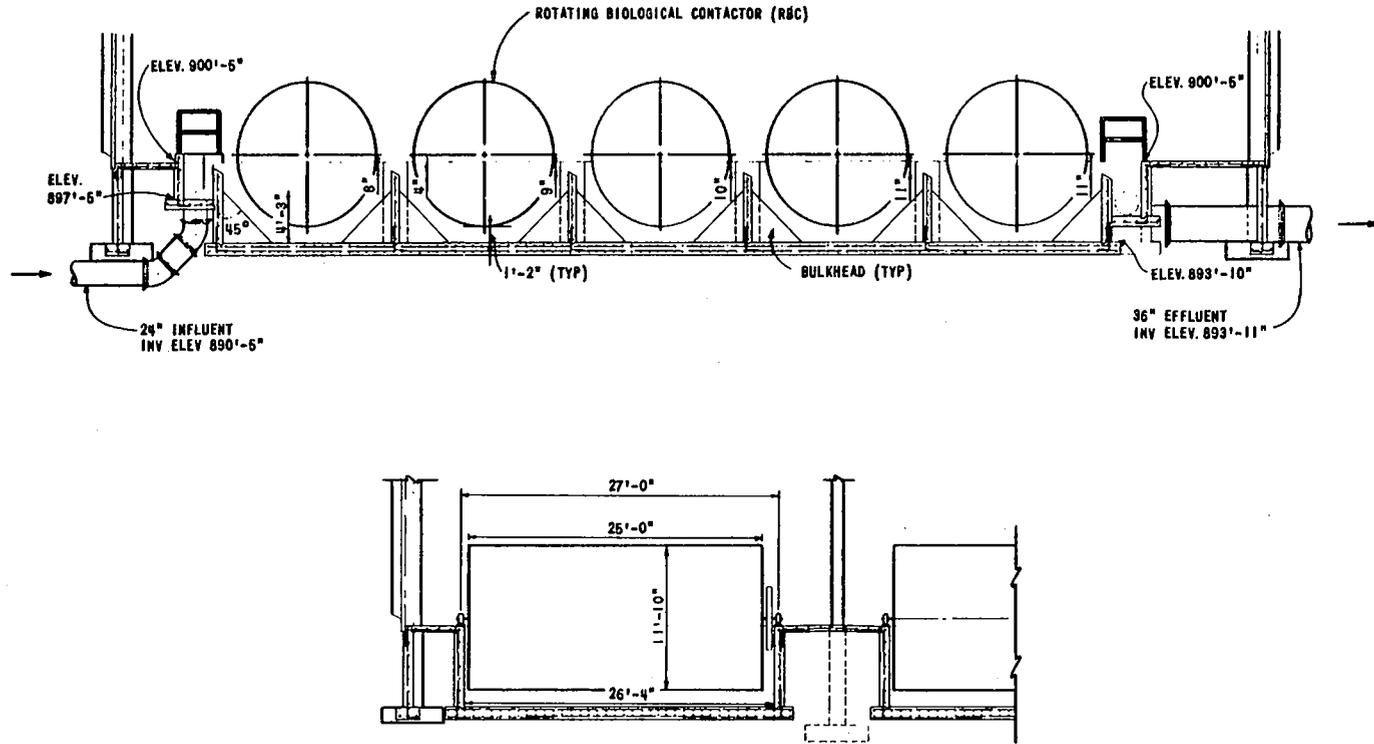


Figure 2. Kirksville RBC Sectional Views

PREVIOUS STUDIES OF THE KIRKSVILLE WWTP

The Kirksville WWTP was included in a nationwide survey of treatment plants (1), conducted to evaluate operational problems and related performance deficiencies of plants located throughout the country. As part of this survey, Kirksville WWTP operating records were reviewed and limited wastewater and treatment process monitoring were performed to assess plant performance. It was determined that the discharge permit limit for BOD had been periodically exceeded and that three of the 20 RBC shafts in the plant had failed structurally prior to the survey in 1978. In addition, extensive corrosion problems were noted within the RBC building.

DuPont and McKinney (2) presented and analyzed operating data for the Kirksville WWTP. They concluded that the plant performed efficiently under normal hydraulic and organic loading conditions, but was susceptible to upset due to organic overloading. As many as six weeks were required for the plant to return to normal operation following an upset.

Decker (3) reported on the operation of the Kirksville WWTP. He noted several plant operating problems, including anaerobic conditions in the first RBC unit in each flow train. He also reported objectionable odors from the treatment plant, and suggested that these might originate from the RBC building, or could be caused by agitating septic influent.

These previous studies documented some of the problems that have been experienced at the Kirksville WWTP. However, they did not provide sufficient data to identify and correct the causes of these problems. It was therefore necessary to investigate the design and performance of the plant in more detail.

PROBLEM INVESTIGATION

Preliminary review of the situation at the Kirksville WWTP indicated that problems with the RBC process were of basically two types. These were (a) periodic inadequate process performance for BOD removal and (b) RBC shaft failures. These two general problem areas are discussed separately below.

Process Analysis

The first step in the process analysis was a review of the design criteria. Process design criteria for RBCs have traditionally been expressed (4) in terms of soluble organic loading rate per unit of media surface area ($\text{lbSBOD}_5/\text{day}/1000 \text{ ft}^2$). Design and current loading conditions for the first-stage RBC units and for the overall RBC process are summarized in Table II.

Design criteria are presented in Figure 3. Also shown is the required performance of the plant, assuming the effluent total BOD_5 (TBOD_5) is twice the soluble BOD_5 (SBOD_5) (subsequent testing indicated that this ratio is reasonable). The process design is consistent with the manufacturer's original overall loading rate criteria. However, based on current design criteria, the plant is undersized. The design overall RBC loading rate of $2.2 \text{ lbSBOD}_5/\text{day}/100 \text{ ft}^2$ at a wastewater temperature of 7°C corresponds to a projected SBOD_5 of approximately 34 mg/l and a TBOD_5 of approximately 78 mg/l .

First-stage loading rate criteria were not established at the time of the facility design. Subsequently, the manufacturer recommended maximum first-stage and overall loading rates of 4.0 and $1.7 \text{ lbSBOD}/\text{day}/100 \text{ ft}^2$, respectively, for nonaerated units. The resulting design and existing first-stage loading rates grossly exceed the recommended limits.

A preliminary troubleshooting investigation was conducted on March 13, 1981. Data were collected over a 24-hour period to establish diurnal fluctuations in TBOD_5 loading. The data collected are shown in Figure 4. These data show that at the time of the investigation, the overall loading was generally within allowable limits, but the first-stage loading exceeded the manufacturer's recommendations much of the time.

A second troubleshooting investigation was conducted in September, 1981. At that time the wastewater temperature was 22°C . With first-stage and overall SBOD_5 loading rates of 6.5 and $1.3 \text{ lbSBOD}_5/\text{day}/1000 \text{ ft}^2$, respectively, the effluent SBOD_5 was 12 mg/l . This is greater than the value (8 mg/l) predicted from the manufacturer's performance curves given in Figure 3. Again, the overall loading was acceptable but the first-stage loading was excessive.

Table II
Kirksville WWTP Organic Loading Conditions

	Flow, mgd	Primary Effluent SBOD ₅ , mg/l	First-Stage RBC lbSBOD ₅ /day-1000 ft ²	All RBCs Organic Loading lbSBOD ₅ /day-1000 ft ²
Existing Conditions (1982)	2.6	150	8.5	1.7
Conditions Corresponding to Design Average Flow and SBOD ₅	3.4	151	8.9	2.2

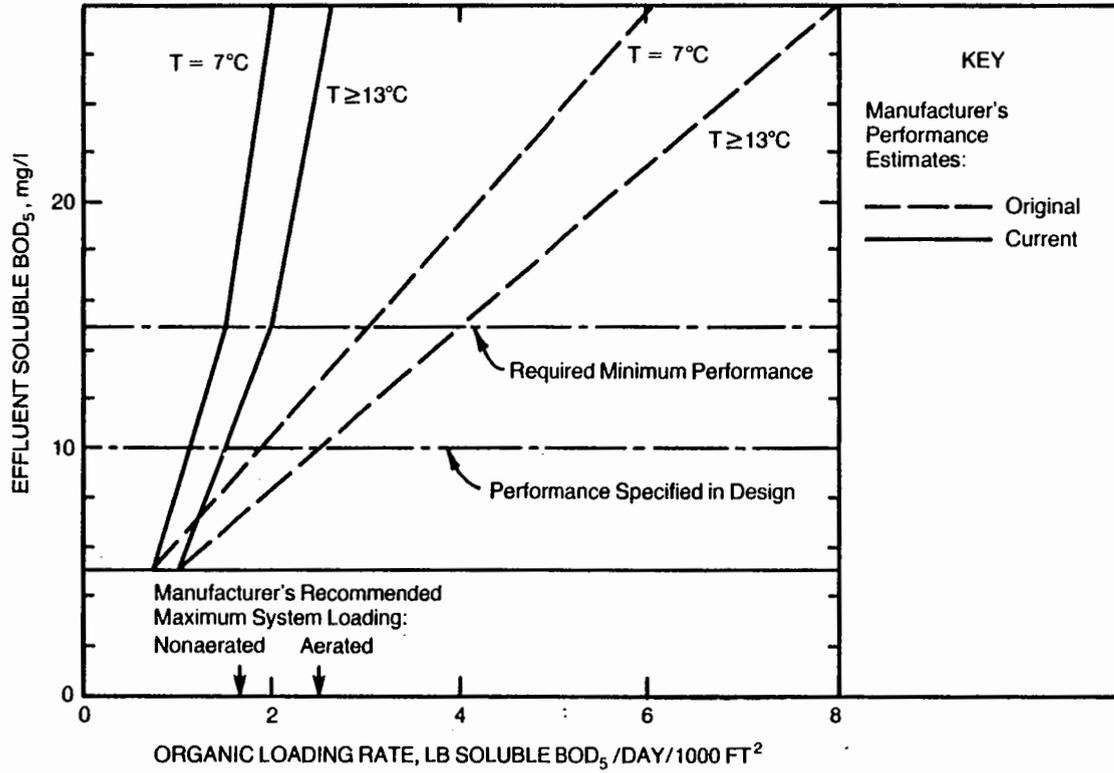


Figure 3. Kirksville RBC Performance Criteria

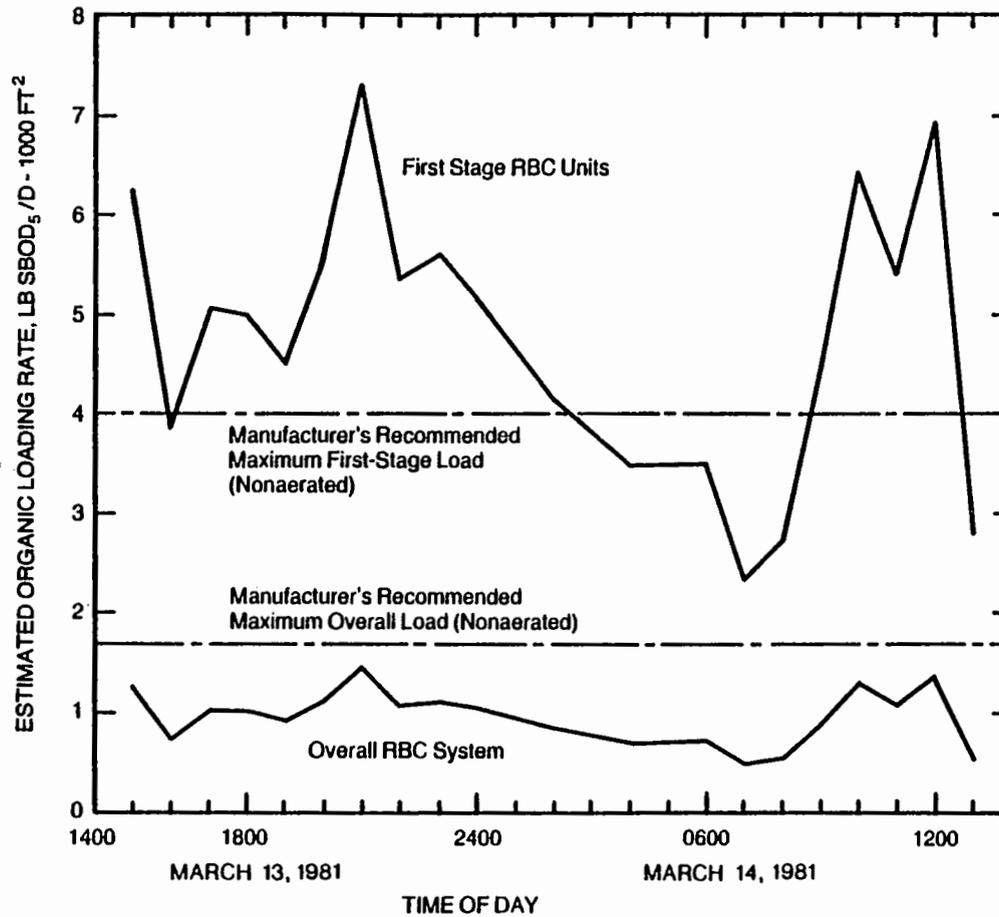


Figure 4. Diurnal Fluctuations in Organic Loading

It was concluded that process modifications would be necessary to improve process performance. Based upon the preceding RBC process analysis, the following process alternatives for upgrading plant performance were selected for analysis:

- I. Increase the number of bays of RBCs.
- II. Reduce the load to the RBCs by primary effluent BOD load shaving.
- III. Increase the RBC load limit by:
 - a. Splitting the flow to the RBCs to establish more first-stage area.
 - b. Removing RBC bulkheads to establish more first-stage area.
 - c. Applying supplemental diffused aeration to the RBCs to increase biomass sloughing and therefore increase the biomass growth rate.

It was determined that first-stage load reduction by flow splitting or removal of existing bulkheads would not satisfy the existing equipment's recommended loading limits at the design year (2000) conditions. A study was designed to evaluate the effects of supplemental aeration on process performance.

Supplemental diffused aeration was provided to the first two units in one "test" train at a rate of approximately 200 cfm. As expected, this resulted in increased biomass sloughing. The system was allowed to equilibrate to the new conditions, and then an intensive study, involving side-by-side tests on the test train and a "control" train, was conducted for a period of approximately 12 days.

Organic loading rates are shown in Figure 5. Overall loads were generally (with some exceptions) within the manufacturer's recommendations, but again first-stage loadings exceeded the manufacturer's recommendations.

Process performance data for the test and control trains are presented in Figure 6 in terms of SBOD₅ concentrations and dissolved oxygen concentrations. These data indicate that aeration substantially reduced SBOD₅ concentrations in

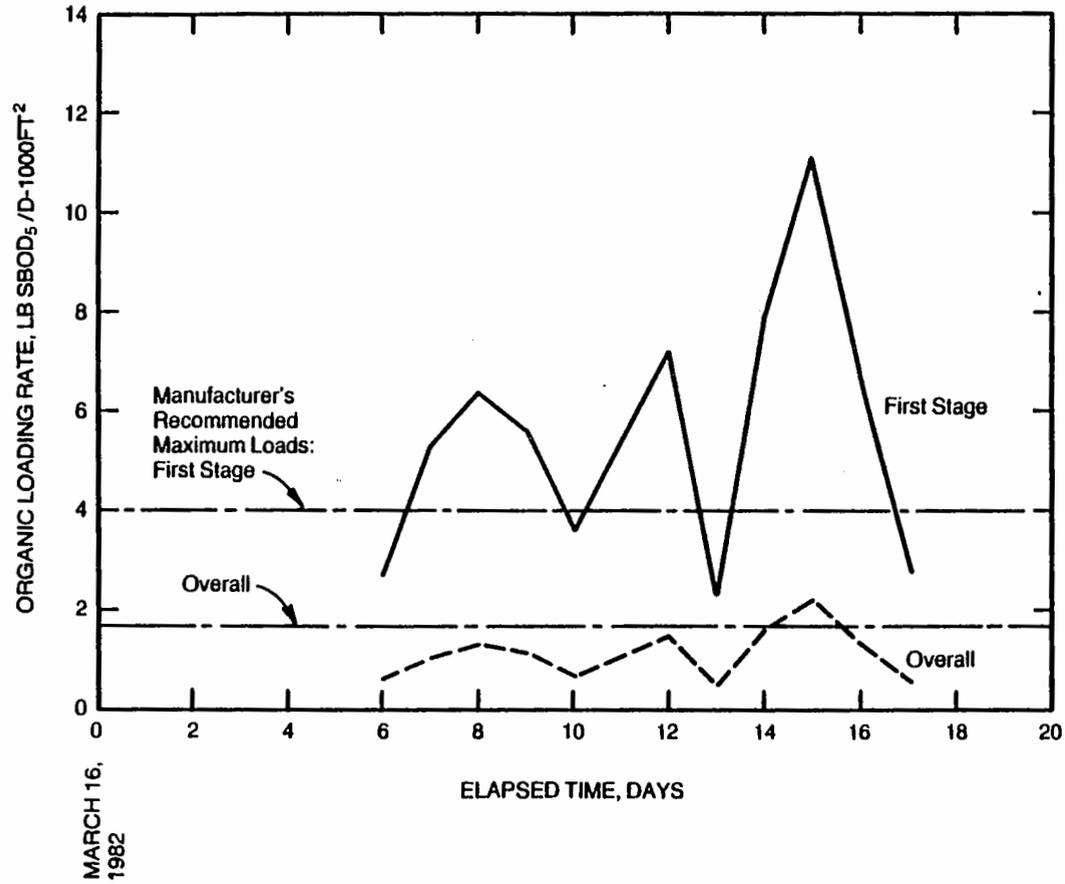


Figure 5. Organic Loading During Aeration Study

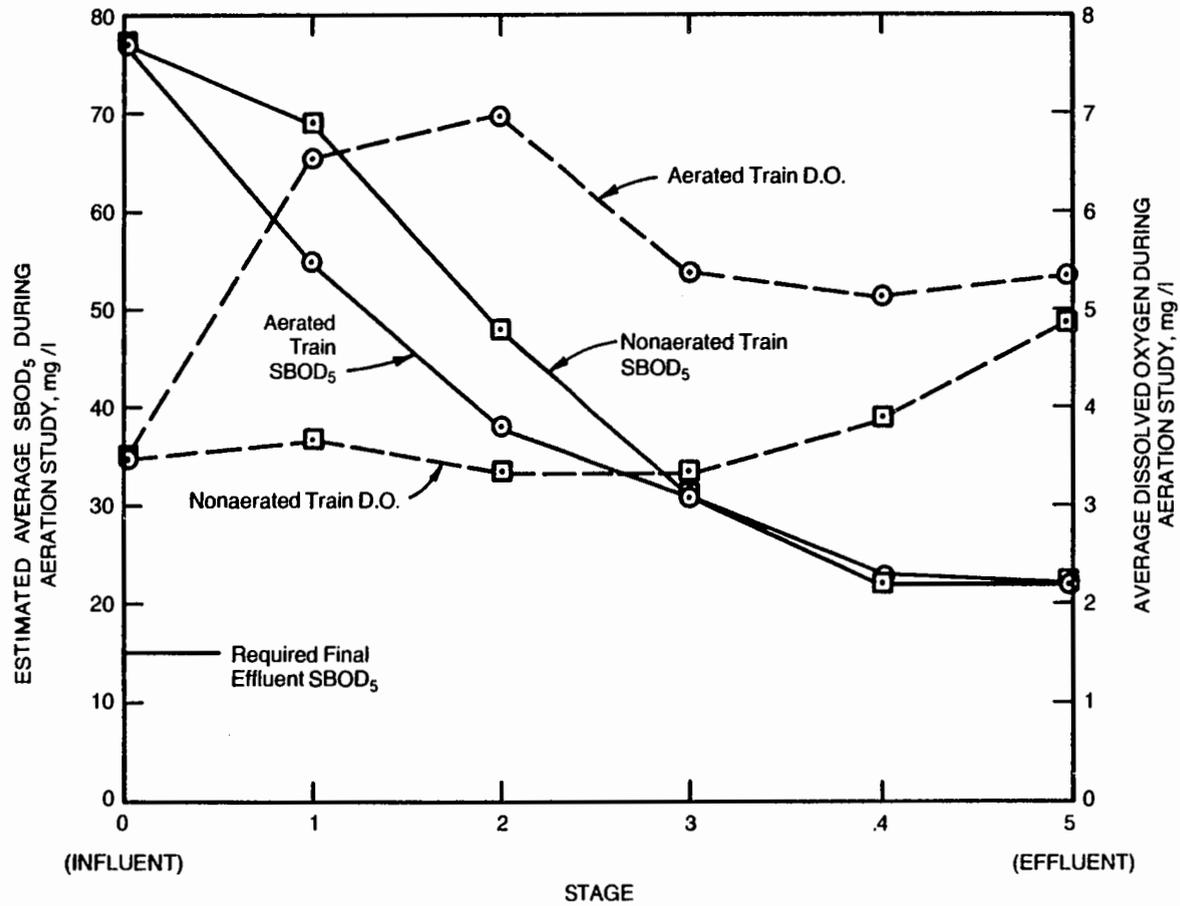


Figure 6. Comparative Process Performance During Aeration Study

the aerated bays, and resulted in higher D.O. values throughout the train. Aeration of the first two bays did not appear to improve the overall performance of the RBC train in terms of SBOD₅ removal.

Oxygen utilization rate data were collected to provide additional insight into the effects of aeration on process performance. Wastewater samples were collected from the fifth bay of both the control and test trains. Average utilization rates for samples collected during the period of the study were 216 and 140 mg D.O./1 hr. g VSS for the test and control trains, respectively. The substantially greater oxygen utilization rate for the test train samples suggests a "healthier" biomass than that in the control train and is an expected result of the reduced mean cell residence time resulting from increased biomass sloughing.

Average volatile suspended solids concentrations in the effluent from the RBC trains for the period of the study were 17 and 40 mg/l for the test and control trains, respectively. The apparently lower volatile suspended solids production in the test train is consistent with the greater oxygen utilization rates observed for samples from the train, and are again indicative of a "healthy" microorganism population.

These results provided strong evidence that supplemental aeration improves the performance of the Kirksville WWTP RBCs in those bays to which aeration is applied. However, the overall performance of the aerated train, when only the first two bays in the train were aerated, was not improved. It is possible that greater differences in performance between the test and control trains would have been observed if the influent dissolved oxygen concentrations had been lower; influent D.O. values averaged 3.5 mg/l during the study. The data presented in Figure 6 suggests that the process was substrate limited in the last three bays in both the test and control trains. With sufficiently low influent D.O. concentrations it is likely that without supplemental aeration the process would become oxygen limited.

Shaft Failure

Problem Review

RBC shaft failure has been the most prominent operating problem at the Kirksville treatment plant since operation began in August of 1976. Two stub end failures and nine mid-span failures occurred within the first five years of operation. Reinforcing rings were installed by the manufacturer

on all 20 shafts in an apparently successful effort to prevent further occurrence of stub end failure. Shafts that failed midspan were replaced with shafts having a greater wall thickness and section modulus and increased strength. None of the replacement shafts has failed to date.

An investigation of shaft failure was necessary to identify probable causes of the problem and to define reasonable loading limits for acceptable future shaft performance. The objective was to develop plant operating criteria to prolong the life of the RBC shafts while satisfying the effluent discharge limitations.

Diagnosis

The locations of all midspan failures are shown on Figure 7. Most failures occurred at approximately one-third the shaft length from the end of the shaft. Inspection of fractured ends indicated probable failure from stress crack propagation. Such cracks could be seen along the surface of the shaft, perpendicular to its axis rotation.

It appears that the RBC shaft failures at Kirksville were the result of metal fatigue. In general, fatigue is associated with repeated discrete load repetitions or, as in the case of an RBC shaft, with complete stress reversals during each operating cycle (revolution). Typically, a material's effective strength declines with the number of cycles of stress reversal if the stress on the material is above a minimum load (the endurance limit). Below this stress level a material can theoretically endure an infinite number of cycles.

There are several design codes for fatigue in steel structures. The RBC industry is not standardized to any particular code. The code (5) recognized by the manufacturer of the Kirksville WWTP RBCs is applicable to the number and magnitude of cyclic stress reversals which can be expected during the typical 20-year planning period for a wastewater treatment facility.

Based on this code and consultation with recognized experts in fatigue analysis (6, 7) it was concluded that the appropriate allowable stress range for the Kirksville RBCs is 10.2 to 13.2 ksi. These values correspond to a design shaft life of 20 years. The lower stress is considered appropriate for long-term average conditions, and the upper range is considered acceptable for short-term (1 to 2 months per year) excursions.

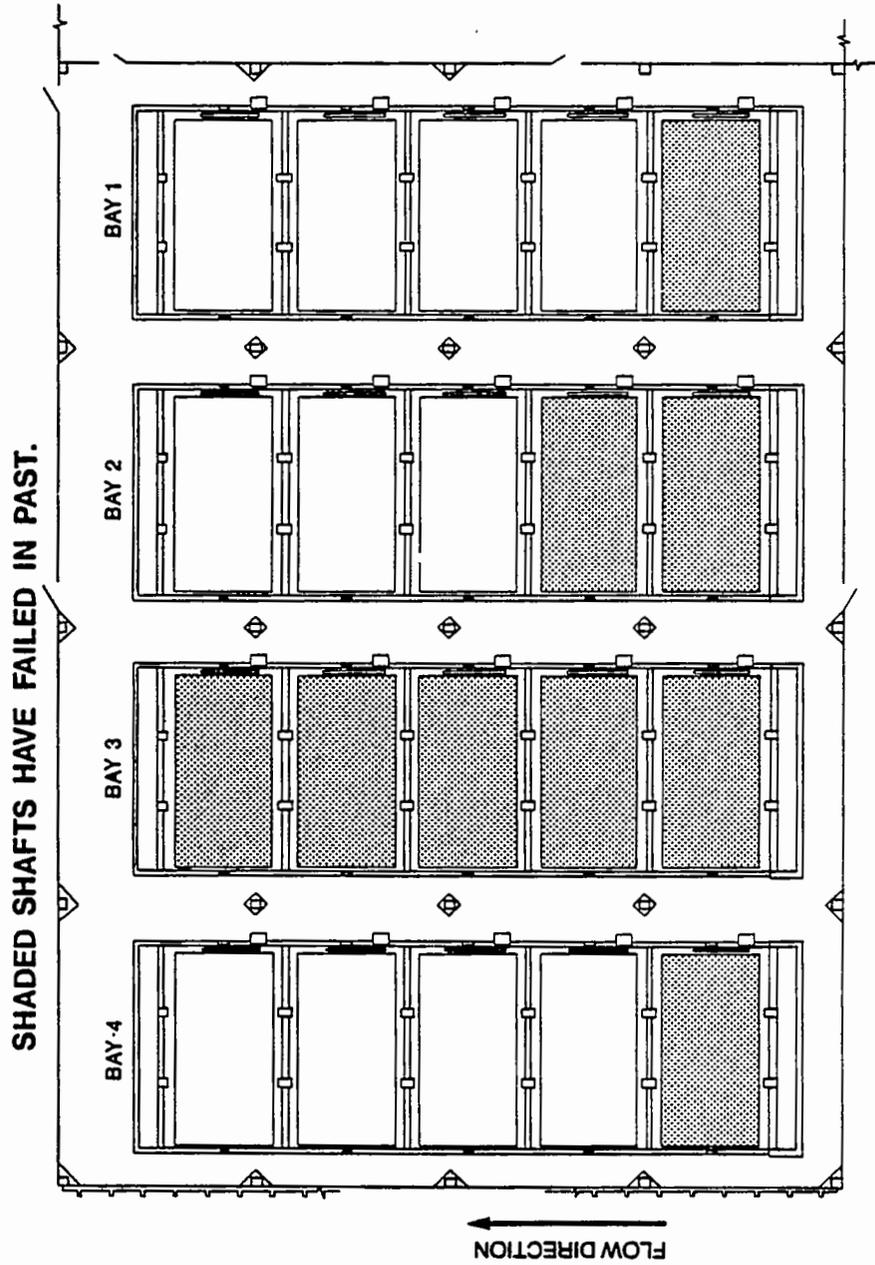


Figure 7. Locations of Shaft Failures

Stress range is presented as a function of measured end load on Figure 8. The relationships for the original and replacement shafts were developed on the basis of bending stress. Assuming the shaft to be a simply supported beam with equal end reactions, maximum bending movement was calculated and divided by the known section modulus to obtain stress. The total cyclic stress range is twice the maximum bending stress because of complete stress reversal. Torsional surface stresses were assumed to be negligible.

Figure 8 indicates that for long-term operation, end loads should not exceed 12,000 or 13,400 lb for the original and replacement shafts, respectively. Short-term excursions to end loads of 15,600 or 17,400 lb for the original and replacement shafts, respectively, are considered allowable.

Testing

A preliminary troubleshooting investigation was conducted in September, 1981 to determine structural loads on the RBC shafts. End load measurements were made on one full bay of RBCs at Kirksville. These and other measurements were subsequently compared to allowable loads indicated on Figure 8. Load measurement was accomplished using a specially fabricated shaft bearing yoke which could be lifted by hydraulic jack along with the entire undriven end of a rotating RBC. Terminal jack pressure measurements were then used to compute load as a function of jack bearing area. The data in Table III were obtained.

Table III
Measured and Allowable RBC Shaft Loads

Unit Number	Shaft Type	Measured End Load lb	Allowable Load, lb	
			Average	Excursion
1	Replacement	19,000	13,400	17,400
2	Replacement	15,240	13,400	17,400
3	Original	13,250	12,000	15,600
4	Original	11,040	12,000	15,600
5	Original	9,280	12,000	15,600

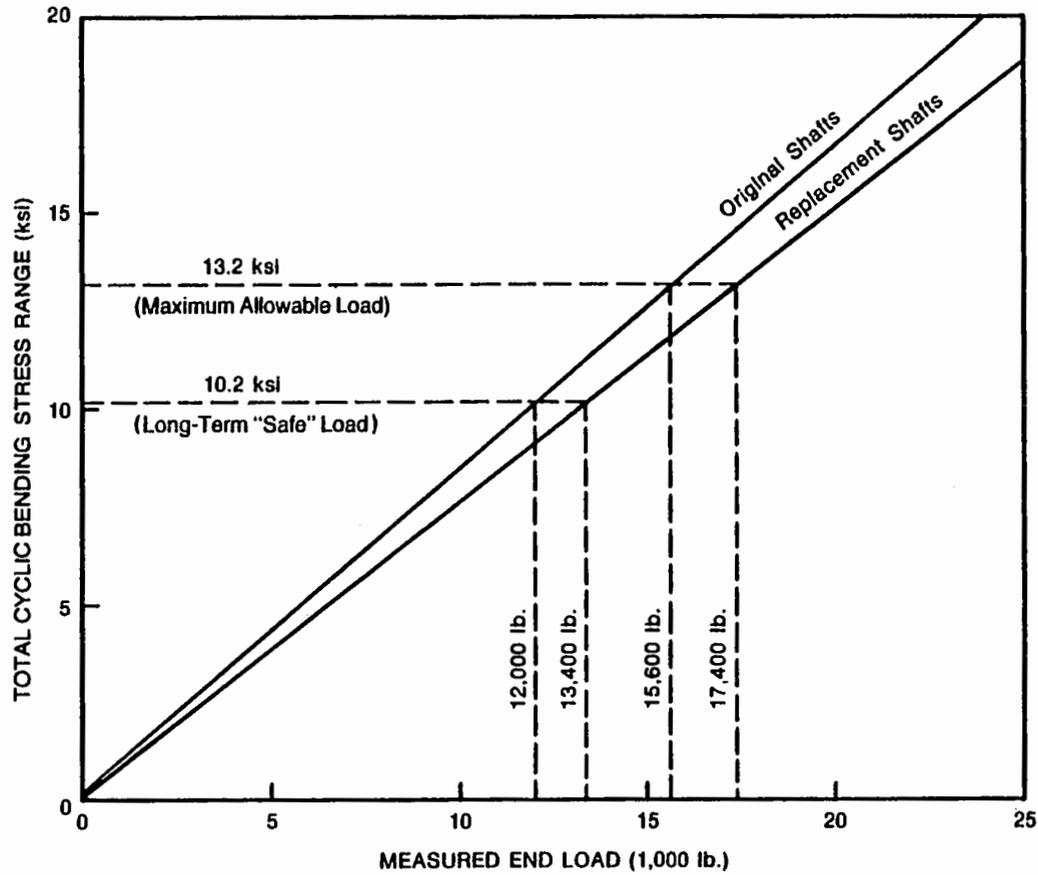


Figure 8. Kirksville RBC Shaft Load-Stress Relationship

These data indicate that at the time of measurement, the first three shafts in each RBC bay were structurally overloaded. The data also provide a plausible explanation for the historical shaft failures at Kirksville. Most of the failures have occurred near the front end of each train. (See Figure 7). Based on the one set of load measurements, the predicted life of the front end shafts would be substantially less than 20 years.

Supplemental aeration was tested as a possible method for reducing shaft structural loads. It was hypothesized that the increased turbulence would result in increased biomass stripping and decreased structural load. The results of this investigation are presented in Figures 9 and 10.

As these data indicate, structural loads on the shafts in the aerated train were less than those on shafts in the control train. All shafts in the aerated train, except stage 3, had loads less than the nominal long-term safe load. It was considered possible that increased biomass stripping on the first two stages resulted in decreased SBOD₅ removal efficiency in these stages, thereby transferring the organic load to the downstream nonaerated shafts. However, data collected to investigate the SBOD₅ removal performance of the process did not indicate this to be the case (see Figure 6).

It was concluded that the increased load reflects "typical" structural loading associated with the organic loading conditions existing at the time of the study. Work is continuing on relating structural loads to organic loads and the effects of aeration.

SUMMARY AND CONCLUSIONS

Analysis of the problems experienced with the RBCs at the Kirksville WWTP indicated that performance problems are associated with organic overloading. It also appears that organic overloading is the principal contributing cause of the large number of shaft failures that have been experienced. This organic overloading results in excessive growth of biomass which is not sloughed, and causes structural overloading of the RBC shafts.

Aeration effectively increases biomass sloughing, reducing the structural loads on the shafts. At the Kirksville WWTP, aeration of the first three bays in each train at 200 cfm should provide adequate biomass sloughing to maintain structural loads at allowable levels, under current and projected future organic loading conditions.

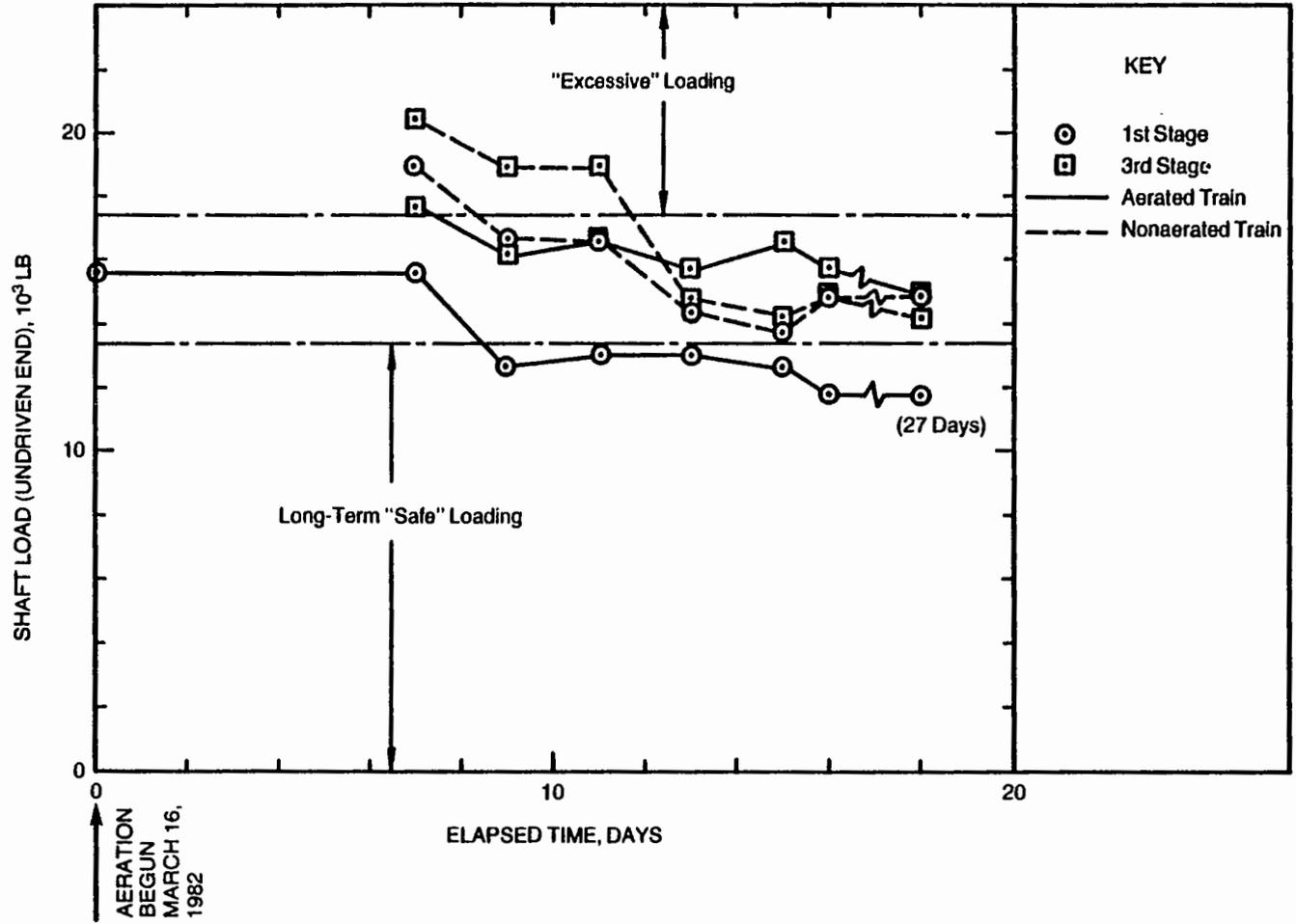


Figure 9. Effects of Aeration on Shaft Load Over Time

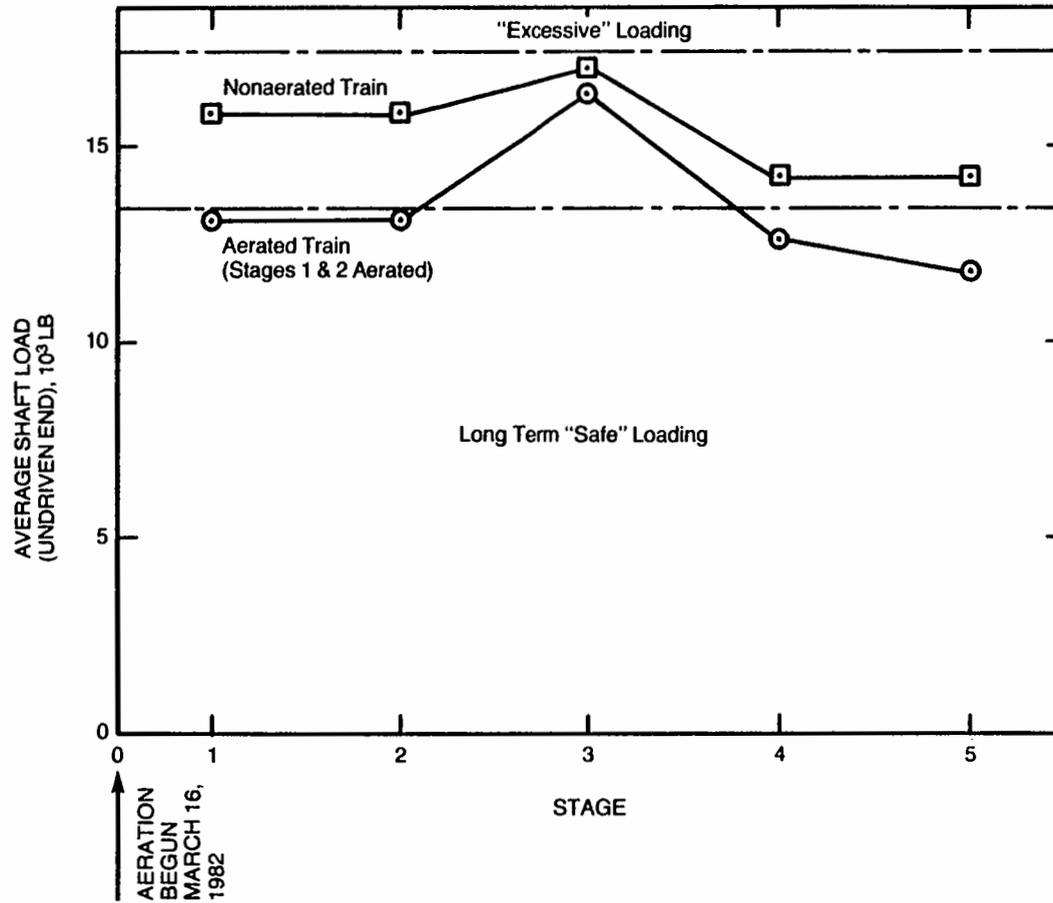


Figure 10. Overall Effects of Aeration on Shaft Load

Some improvement in process performance for SBOD₅ removal appeared to result from aeration. However, it was concluded that at the design organic loading the WWTP will likely not meet its performance objective even with reasonable supplemental aeration. Supplemental aeration, combined with reductions in loadings to levels recommended by the manufacturer, will be required to provide adequate wastewater treatment. An important conclusion from this study was that without supplemental aeration, structural failures of the RBC shafts are likely to continue.

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STRUCTURAL ENGINEERING OF PLASTICS MEDIA
FOR WASTEWATER TREATMENT BY FIXED FILM REACTORS

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INTRODUCTION

A balance between process efficiency and structural integrity is the prime objective in designing a large scale fixed film reactor for wastewater treatment. Various criteria must be considered before the proper biological oxidation media can be selected to achieve an optimum design.

This paper examines the criteria for selecting plastic media, its performance standards, and its properties, such as its modulus of elasticity at the maximum working temperatures it will withstand; the ability of the media to retain these properties when exposed to the deleterious effects of chemicals, physical impact, and ultraviolet light; and the media's ability to remain efficient as biota growth accumulates on its surfaces.

Performance standards for determining the maximum load plastic biofilter modules can support are also discussed. These standards were developed to provide a media which has structural strength to remain functional for the life of the process.

Test procedures for predicting load bearing capabilities of the plastic media are developed. A test is described that simulates the physical capabilities of the plastic media modules when they are supporting biota growth. This data is important in establishing the structural design of the modules.

Reference is made to recent attempts to develop in-service load measurements by installing load cells in a tower during construction.

BIOFILTER INSTALLATION

A fixed film reactor for bio-oxidation of wastewater consists of a containment vessel, a fluid distribution system, surface system for biota (biomass) growth, support for the media (base), a process effluent collection system, and an air supply. (See Figure 1.)

In early biofilter installations, the surface for biota growth (biological oxidation media) was provided by random rock. During the past decade, however, fabricated modules have grown in popularity. These modules offer many advantages. They can be stacked, are self-supporting, and generally do not impose a load on the containment walls. Other advantages of fabricated modules are that they do not plug as does rock, and since they are much lighter than rock (one-thirteenth the weight of the same volume of rock), the cost for their support base within the containment vessel is less than that required for rock.

The support base for the plastic media modules must be designed so that it does not interfere with the flow of process effluent nor with the air flowing up through it to the media modules. The base for most plastic media is made up of a series of precast, narrow, slotted dense concrete beams. (See Figure 2.) The beams are placed at 2-ft. centers and are supported by cast-in-place concrete or block posts that are placed as required for structural loadings. Each module on the bottom layer is supported at three places (at both ends and at the center) to establish a stable base for succeeding layers. The use of the beam and post design results in substantial areas of free air space below the media to minimize obstruction to air flow. Random dumpfill and some modular media that lack structural integrity require additional support of grating to span the beams.

DESIGN OF FILTER MEDIA MODULES AND THEIR INSTALLATION

Plastic media modules for waste treatment appear in many shapes, and are fabricated using a variety of techniques. Three overriding considerations dictate the final form a media will assume:

- (a) It must be functional in the process it serves
- (b) It must have the structural strength to remain functional for the life of the process
- (c) It must be economical to fabricate

Plastic media, of the type discussed in this paper, is fabricated from alternating flat and corrugated PVC sheets

that are generally two feet wide and four feet long. These alternating sheets are bonded together to fabricate a self-supporting module that is two feet wide, four feet long, and two feet deep. This construction (See Figure 3.) provides columnar strength, permitting biofilters to be stacked to a 40 foot height without intermediate supports.

One module, which has an ideal shape and fabrication technique is shown in Figure 4. Its straight column design gives maximum strength. A zig-zag pattern in formed corrugated sheets prevents straight fall-through of the wastewater and insures excellent oxygen transfer. While breaking a column by adding waves and shapes will enhance its process efficiency, it could reduce the structural strength and make fabrication more difficult.

Plastic modules have sufficient structural integrity so that when they are placed directly on the slotted concrete beam supports, they will be strong enough to resist the compressive forces imposed at module-support interfaces. These forces are a result of the module weight, the weight of the biota accumulation and the transient hydraulics passing through the biofilter.

During installation the bottom layer of modules is arranged in rows that run at right angles to the support beams. (See Figure 5.) Subsequent layers are arranged in a herringbone pattern, at a right angle to the row of modules preceding and supporting them. (See Figure 6.) This pattern effectively ties the stack of modules together so that the entire structure is self-supporting.

Since the media imposes no load on the containment wall, the wall need be only strong enough to support itself, for local windloading conditions, and act as a sight shield. The wall should be opaque to shield the plastic media from U.V. radiation, and it should be corrosion and weather resistant.

ESTABLISHING THE LIVE LOAD

Since the plastic modules range in weight from two to five pounds per cubic foot, they do not exert a significant load on themselves. Thus, the structural requirements for self-support is minimal. The weight of biota accumulation per cubic foot of module, however, increases the weight to as much as 13 pounds per cubic foot (based on 27-30 ft²/ft³ media).

This live load in structural design formulas of 13 lbs./ft³ can be calculated by summing the following:

1. Maximum Biota Growth
2. Weight Media
3. Weight of Transient Water

In tests conducted under laboratory conditions it was determined that the weight of the bacterial film that accumulates can be as high as 12.9 pounds per cubic foot of media.¹ In tests conducted by BFGoodrich in 1975, samples were taken from an industrial waste application. Results of this sampling indicated the biota accumulated weighed 9.2 pounds per cubic foot. On-going measurements are currently being conducted in a municipal waste treatment tower in which strain gauges were placed under the beam support structure during installation. These load cells should make it possible to compare the weight of the stacked media at the start-up of the process with the weight of the media after biota growth has accumulated and stabilized. The rate at which biota accumulates also can be studied through this instrumentation. Preliminary data indicates that the average weight of the accumulated media is 10 lbs./ft³.

PROPERTIES OF PLASTIC MEDIA

Synthetic polymers have good resistance to water, oxygen, bacterial action, weak acids, and alkali. Therefore, the selection of a synthetic polymer as the material of construction for biofilter media was an obvious choice. The type of synthetic polymer ultimately selected is contingent upon its cost, weather resistance, temperature resistance, and resistance to organic chemicals that may be present in the waste. Since most biofilters operate below 135°F., unplasticized PVC has shown to be the most cost-effective for this application. For this reason it is the most commonly used in fabricating waste treatment media.

One of the first installations to use PVC plastic media was completed in 1961.² The Rome Kraft Company, Division of the Mead Corporation installed 100,000 cubic feet of PVC media to treat effluent from their paper operation in Rome, Georgia. PVC was selected as the material for the media after research had been conducted on a number of different materials. In 1969, a sample of the PVC media was withdrawn and its physical properties tested. There was no apparent change in the material.

Typical properties of the PVC sheet which is used to fabricate the modules are:

Tensile Strength-psi	ASTM D-638	4,000-8,000
Modulus of Elasticity-psi	ASTM D-747	325,000-400,000
Impact Strength-Ft.Lb. Per in. of Notch	ASTM D-256	1.0
Heat Distortion Temperature @ 264 psi	ASTM D-648	155°-160°F
Maximum Recommended Service Temperature-°F		135°F
Specific Gravity	ASTM D-792	1.40-1.60

Because the structural demand on the media is compressive in nature, the Modulus of Elasticity is critical to the load bearing capability of the media.

PVC sheet, from which modules are fabricated, is a thermoplastic, and its properties are affected by the ingredients that go into its formulation. For example, certain stabilizers can act as plasticizers which can adversely affect the load bearing capability of a module at higher temperatures. Figure 7. shows how the modulus of elasticity of a properly formulated PVC is affected by temperature. Additives also influence PVC toughness and its resistance to weather and chemicals.

Another property affected by additives and temperatures is the cold flow in polymers which are subjected to long-term stresses. Thermoplastics that are subjected to a constant load will deform quickly to a certain point that can be predicted by its stress strain modulus. Beyond that point, the thermoplastic will continue to deform indefinitely at a slower rate. This slower deformation rate, or cold flow, is called "plastic creep"³.

Creep rate varies directly with:

1. Load
2. Temperature
3. Ductility

The creep rate in rigid PVC is also affected by its compounding and its chemical environment. Both of these alter the creep rate since they affect both ductility and the stress strain modulus.

The creep rate of PVC increases when:

1. Molecular weight decreases
2. Plasticizer content increases
3. Lubricant content increases, and/or
4. Environment contains water and/or solvents

The creep rate of PVC decreases as:

1. Slow oxidation or aging takes place
2. Filler content increases.

The rigid PVC material is designed to have a low creep rate and good serviceability; the creep rate value of 2.5 is used for design. This value is compensated for the softening effects of water and traces of solvent that are present in wastewater as temperatures reach and exceed 100°F.

The 2.5 value for creep rate is based on long-term experience and empirical data acquired during the 20 years that PVC compounding and engineering design for waste treatment media have been practiced.

If we take a square foot cross-section of a module with a .015 in. sheet thickness, the area supporting the load is 2.7 square inches. The maximum load imposed on the .015 in. sheet occurs at the 14 ft. depth as shown in Figure 8., where the design load is 218 lbs. Dividing the design load of 218 lbs. by the support area of 2.7 square inches yields a stress of 81 psi. The maximum design stress would occur at the 28 foot level. At that point the media is .023" thick and the load is 437 lbs. which yields a design stress of 106 psi.

At these relatively low stress values one would expect a low plastic creep factor in the range of 1-2.25. (Published creep data on PVC compounds start at stress levels of 1000 psi or higher.) However, because of the affect that the environment has on the modulus of elasticity of the PVC a plastic creep factor of 2.5 is used.

ESTABLISHING LOAD BEARING CAPACITY

Once a module of a certain design is selected, it is necessary to establish its load bearing capability. The ability of a module to support a load varies with its sheet thickness (after forming) and the amplitude of the corrugations as shown in Figure 9.

The load bearing capability of a module is called its Load Rating and is defined as that load at which the module will deflect 1%. The test procedure for determining the load rating is detailed in Appendix A. Tests are performed at 75°F. and are adjusted to 105°F. by using the graph in Figure 7. Figure 10 shows this adjustment. A series of compression tests is conducted on like modules and the results averaged to obtain a load rating.

A variation of this test is used for the base layer of media. Since the load bearing area for these is less at the point where the media rests on the support beams, the bottom layer of media must have a higher rating.

Since the load on a module located near the bottom is greater than on one near the top, the engineer must take into account the total weight of all the modules stacked on top of any one module. As was stated earlier, the live load of a module is 13 pounds per cubic foot per foot (the combined weights of the media, biomass and transient hydraulics). For design purposes, however, the live load is adjusted by adding a 20% safety factor and a "plastic creep" factor of 2.5. Thus, the specified test load would be:

$$(13 \text{ lbs./ft.}^3 \times 1 \text{ ft.} \times 1.2 \times 2.5 = 39 \text{ lbs./ft.}^3 \text{ft.})$$

To determine the test load for a module at a specific location in the biofilter, the depth is multiplied by the specified test load.

The following example shows the tabulation of calculations made for a theoretical biofilter, with a media depth of 20 ft. and a specified test load of 39 lbs./ft.³/ft., at an operating temperature of 105°F. It is also specified that the minimum test load rating of any module be 500 lbs./ft.² and the top and bottom modules are specified at 1000 lbs./ft.².

<u>Media Depth</u>	<u>Spec. Test Load</u> lbs./sq.ft.	<u>Low Capability</u> <u>PSF</u>		<u>Sheet Thickness</u>
		75°F	105°F	
1. Top Layer 2'	1000	1152	1000	.023"
2nd Layer 4'	156	712	618	.015"
3rd Layer 6'	234	712	618	.015"
4th Layer 8'	312	712	618	.015"
2. 5th Layer 10'	390	712	618	.015"
6th Layer 12'	468	712	618	.015"
7th Layer 14'	546	712	618	.015"
3. 8th Layer 16'	624	892	774	.017"
9th Layer 18'	702	892	774	.017"
4. 10th Layer 20'	1000	1000	1200	.023"

Following the principles stated above, the structural performance of PVC Biofilter media produced by BFGoodrich has been proven over the past 20 years. With approximately 14 million cubic feet supplied no structural failures have occurred.

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Appendix A

TEST FOR COMPRESSIVE PROPERTIES OF PLASTICS MEDIA FOR WASTE WATER TREATMENT

Scope:

This procedure covers a method of measuring load bearing qualities of rectangular modules of plastics media. Modules are placed in a press in an arrangement simulating their relationship when in service. Load-deformation curves provide a tool for expressing load values for specific deformation levels. This test relates only to the load-deformation behavior at media-to-media Contact.

Significance:

Yield points on the load-deformation curve indicates the load which will collapse the media. At some load less than the yield point, a deformation limit can be set which will represent the maximum which can be tolerated by the process in which the plastic media is to be used.

Before application in engineering design, the load value determined by this procedure must be adjusted to correct for differences in the service environment and the test environment. Allowances must be made for change in modulus of the plastic material with change in temperature. Allowance must also be made for plastic creep.

Apparatus:

1. Test stand with a rigid top and bottom which will not deflect when a load is applied. Space between the top and bottom should be large enough to accommodate two full size (2' high) modules plus air pillow and four 3/4" thick pieces of plywood. Width and length of test stand must accommodate full size modules (2' wide X 4' long).
2. Air pillow to apply load made to a 24" X 24" contact area when inflated.
3. Mercury manometer which reads in inches of water. Air pressure to inflatable bag should be connected to one side of manometer with the other side of the "U" tube opened to atmosphere.

4. Air source - two regulators and needle valve for bleed line.
5. One or two 30 - 40 gallon surge tanks.
6. Three ring stands with holding clamps.
7. Three dial indicators which will read 1.000" deflection.
8. Three 1 1/2" X 8" X 1/8" metal strips.
9. Four pieces of 3/4" plywood cut to 26" X 26" with a 2" X 2" square notch cut in one corner. Plywood pieces must not be warped.

Test Specimen:

Size: 24" high X 24" wide X 48" long

Number Required: Two per test

Conditioning: The specimen must be clean and dry. The specimen shall be stored at test temperature for at least two hours before testing.

Procedure: (See Figure 11.)

1. Place air pillow in center of bottom of the test stand.
2. On top of the air pillow, place two of the plywood boards so the notched corners are at the same point as the air valve. Center the boards on the air pillow.
3. Weigh two test modules and record the weights. Place a full size module in the test stand parallel with the framework. Center the module on the 26" X 26" plywood board. Place a second full size module of the same construction (guage and pair number) on top of the first pack and 90° to the bottom module. The top module should be centered on the bottom module and the 26" X 26" plywood board.
4. Place two plywood boards on top of the two modules and be sure boards are centered over the modules and the bottom plywood boards. If there is additional space between the top of the board and top of the test stand, fill in the space with additional pieces of unwarped plywood or 2 X 4's until fit is snug. Measure the actual interface area of the two modules and record.
5. Slide the 1 1/2" X 8" X 1/8" metal strips between the bottom module and plywood board as shown in figure 12.

6. Place the dial indicators in the holding clamps and place the plunger of the dial so it is barely touching the metal strip.
7. Attach air line to bag (air system shown in Figure 13).
8. Fill surge tank to 20 - 25 psig using air regulator ①.
9. Using air regulator ②, apply 30 inches of water pressure to the air pillow. This will snug-up the system and take out any slack. Let stand for two minutes and adjust dial indicators to zero.
10. After zeroing dial indicators, increase the pressure in increments of 30 inches of water every 20 seconds. Read and record dial readings on the three dial indicators after each pressure increase.
11. When specified load and/or % deflection has been reached, hold the load for 60 seconds. Read and record dial readings on all three dial indicators, average and record as the settle point.
12. If the testing is to increase at 30 inches of water every twenty seconds and record dial readings until the rate of deformation sharply rises. The load at the point of rise is the yield point. (For quality control purposes, the test need not be continued beyond the deformation limit or the maximum load rating agreed upon between vendor and customer.)

Calculations:

1. Convert the pressure reading from inches of water to pounds per square foot.
2. Average the three dial readings for each load level.
3. Construct a load deformation curve.
4. Using a straight edge, draw a line between the zero load zero deflection point and the settle point.
5. To determine the load at 1% deflection, read the load off the graph at .48" deflection.

Reports:

1. The load in pounds per square foot at 1% deformation.
2. The percent deformation at a designated test load.
3. The yield point in pounds per square foot - if required.
4. The test temperature.

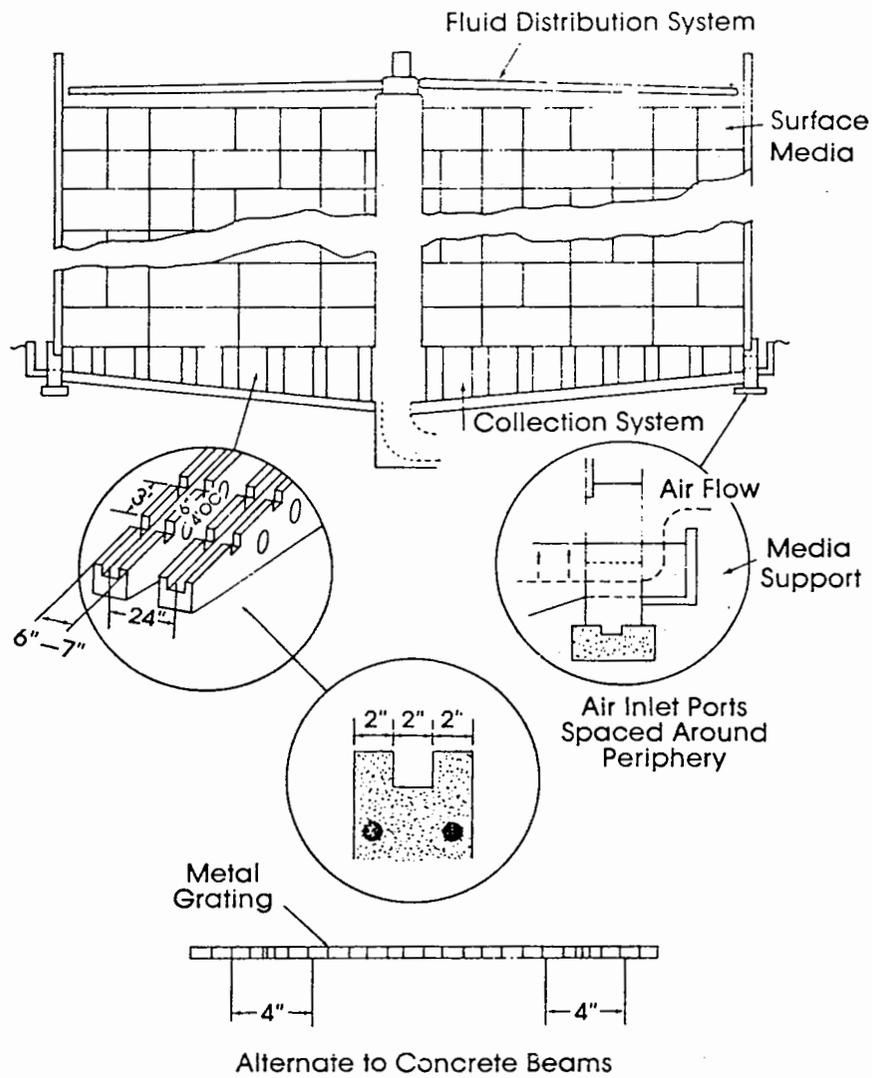


Figure 1. Fixed-film Reactor for Bio-oxidation.

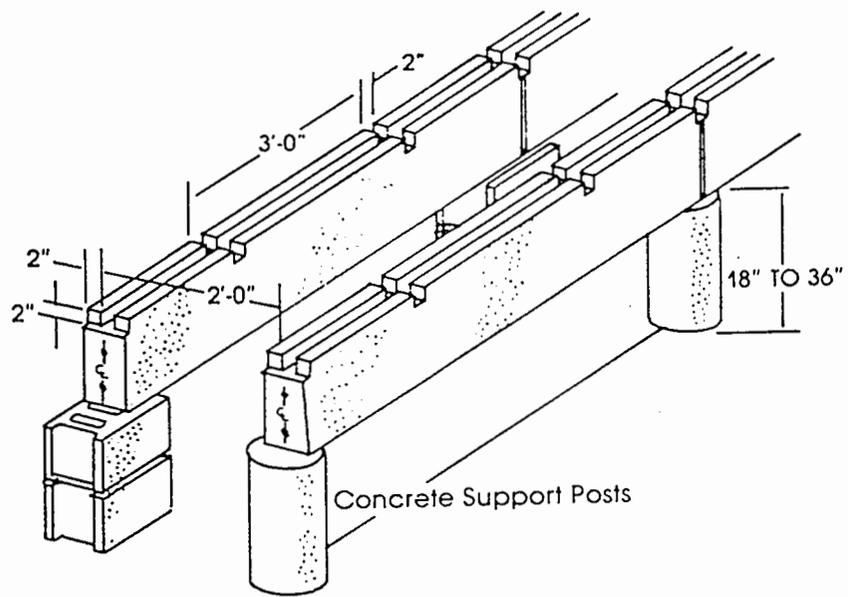


Figure 2. Media Support.

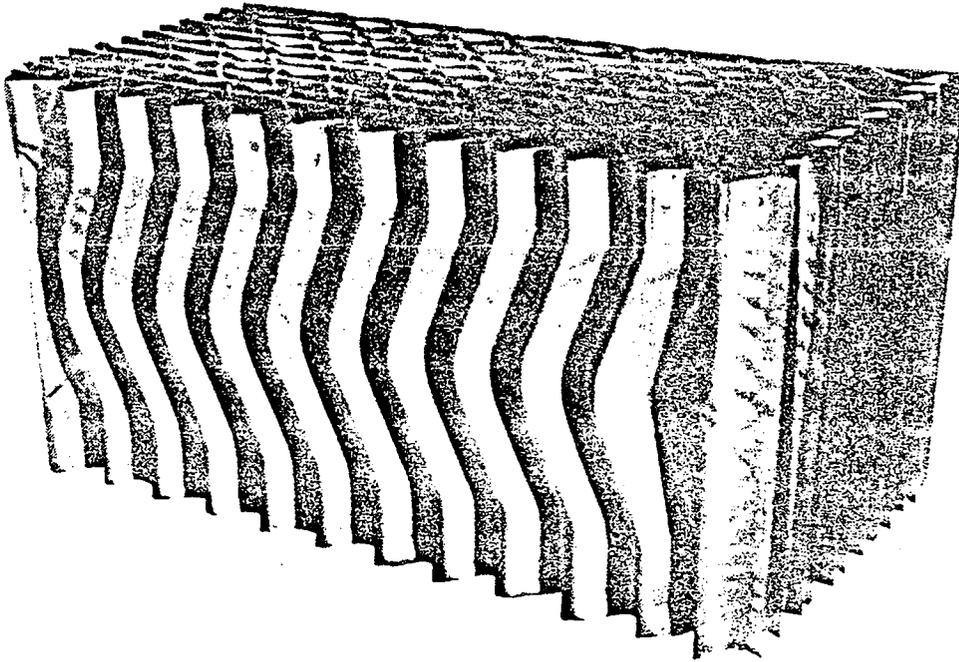
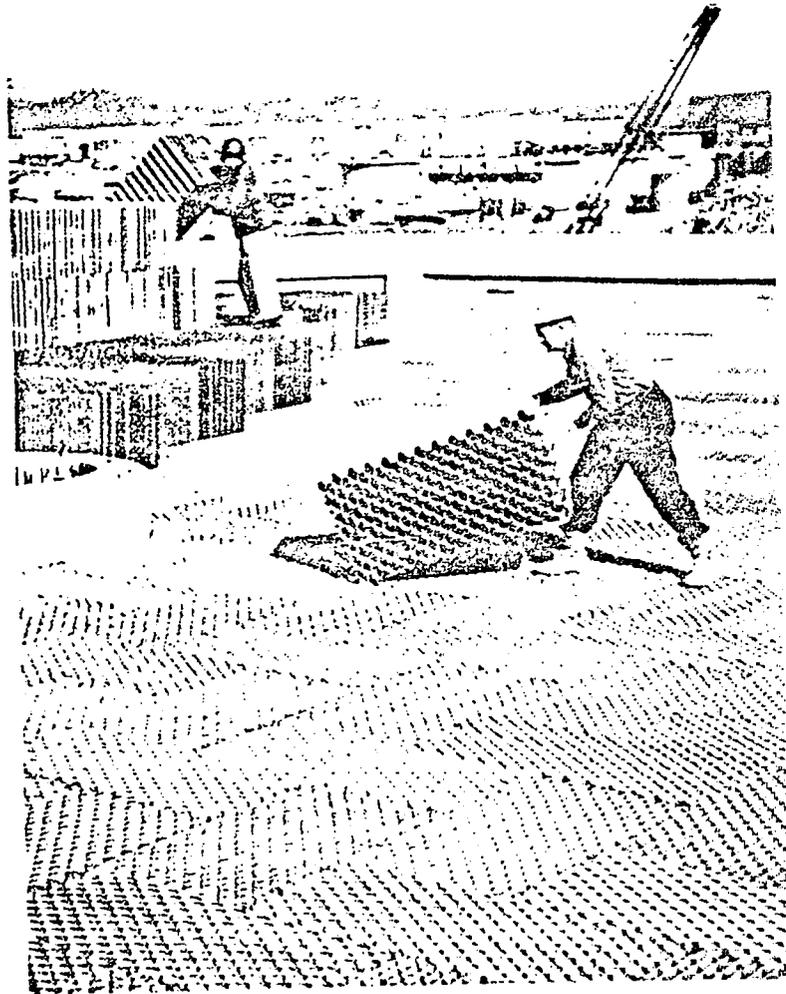


Figure 3. Flanged Plastic Media



**Figure 4. Media Installation:
Maximum-Strength Pattern.**

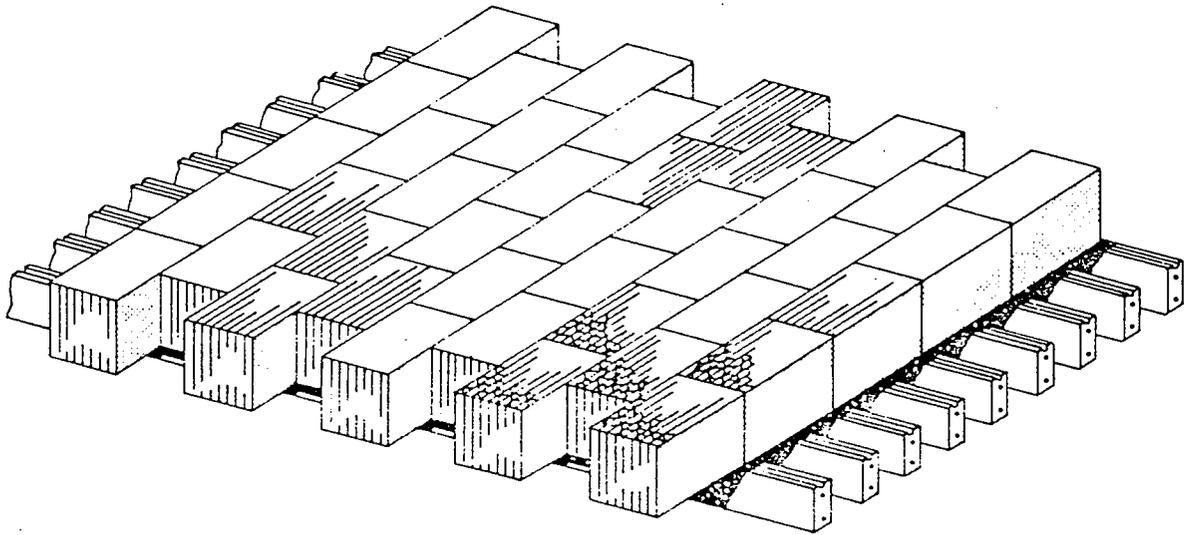


Figure 5. Base Layer Arrangement.

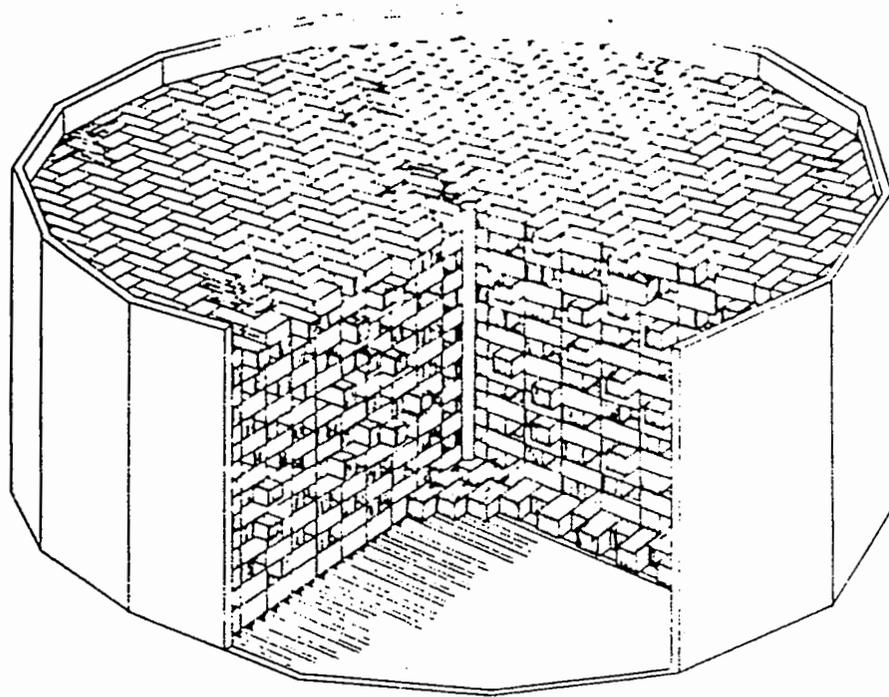


Figure 6. Packing Pattern.

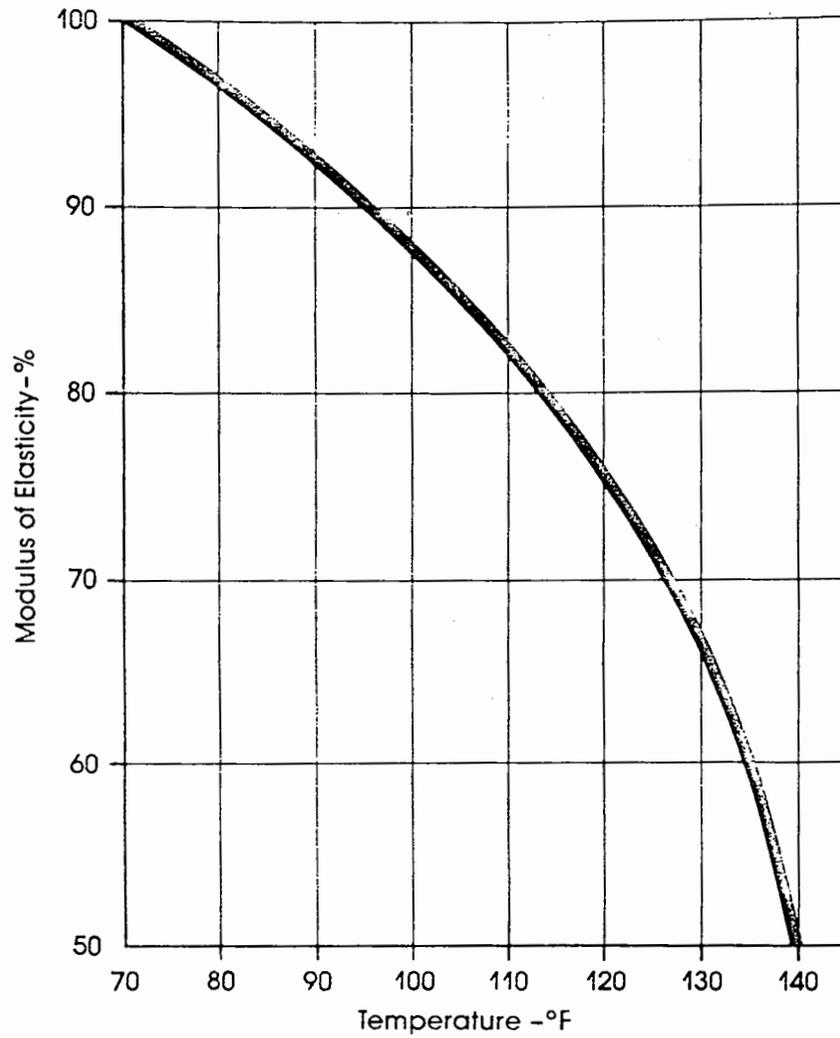


Figure 7. Rigid PVC: Modulus vs. Temperature.

	Media Depth in Tower (Ft.)	Panel Thickness Inches	Live Weight Lbs.	×	Safety Factor 1.2	=	Design Load Lbs.
Top	2'	.023	26				31.2
	4'	.015	52				62.4
	6'	.015	78				93.6
	8'	.015	104				124.8
	10'	.015	130				156
	12'	.015	156				187.2
	14'	.015	182				218.4
	16'	.017	208				249.6
	18'	.017	234				280.8
	20'	.019	260				312
	22'	.019	286				343.2
	24'	.019	312				374.4
	26'	.023	338				405.6
	28'	.023	364				436.6
	30'	.035	390				468

Figure 8. Design Load versus Media Depth

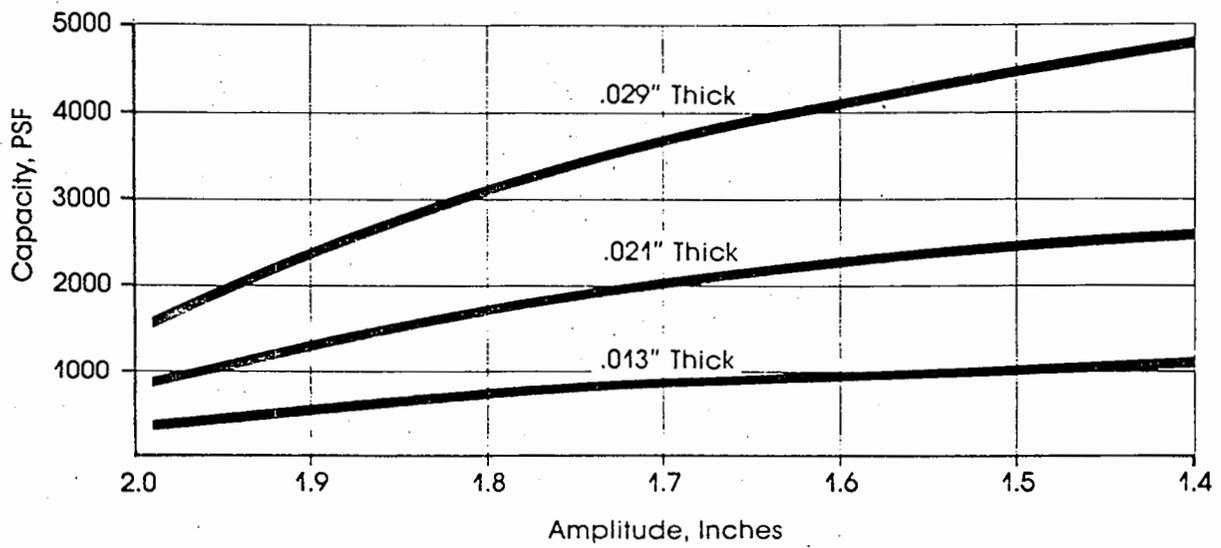


Figure 9. Membrane Amplitude, Inches vs. Load-bearing Capacity.

Thickness Inches	PSF Load Rating 75°F	PSF Load Rating 105°F	PSF Design Rating 105°F
.015	712	618	247
.017	892	774	309
.019	1092	947	378
.021	1312	1138	455
.023	1552	1347	538
.025	1812	1572	628
.027	2092	1815	726
.029	2392	2076	830
.035	3412	2961	1184
.035 Beam	2000	1736	694

The load rating at 75°F is the average result of many instantaneous load tests at 1% deformation. The load rating at 105°F is calculated from the ratio of the Modulus of Elasticity illustrated in Figure 7. The design rating is the load rating at 105°F divided by a plastic creep factor of 2.5.

Figure 10.

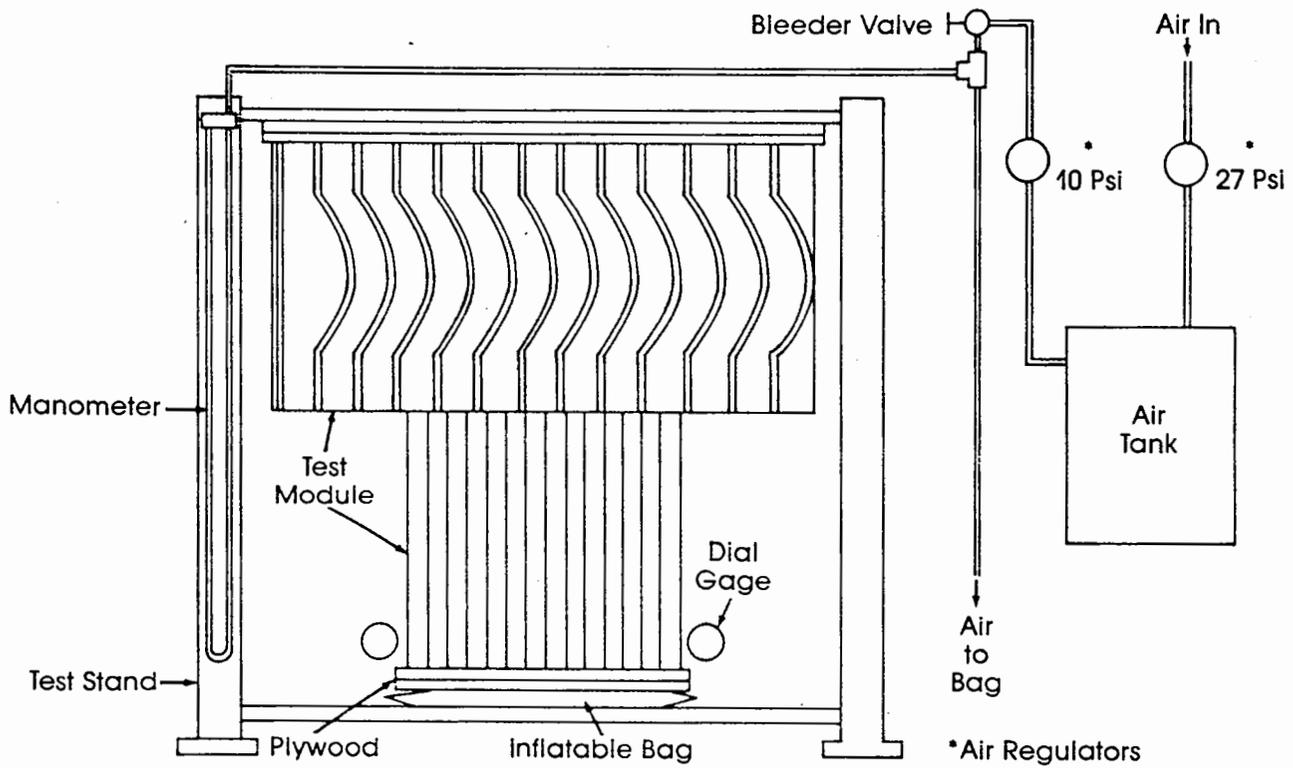


Figure 11. Set-up in Test Stand.

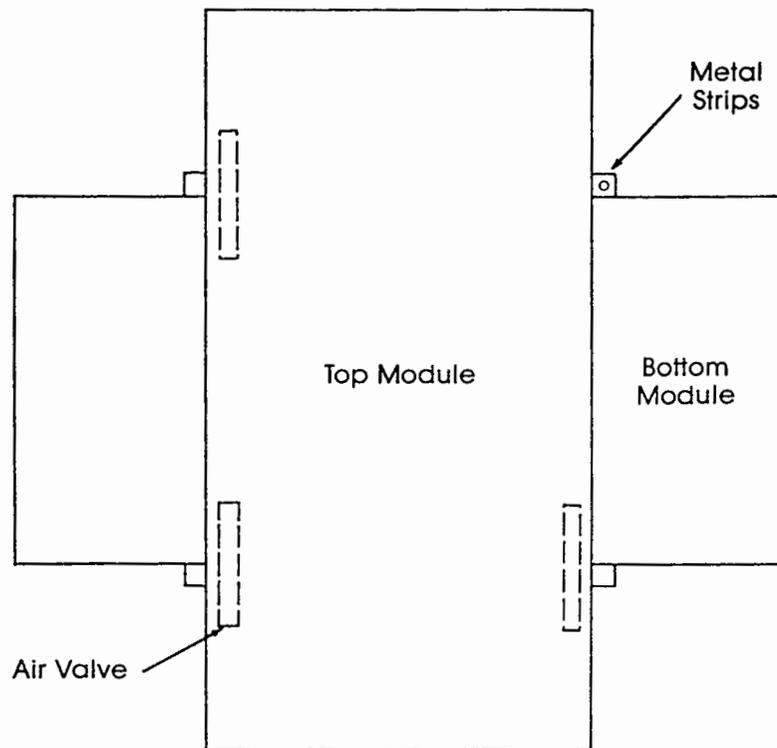


Figure 12. Arrangement of Test Modules and Metal Strips For Deflection Test.

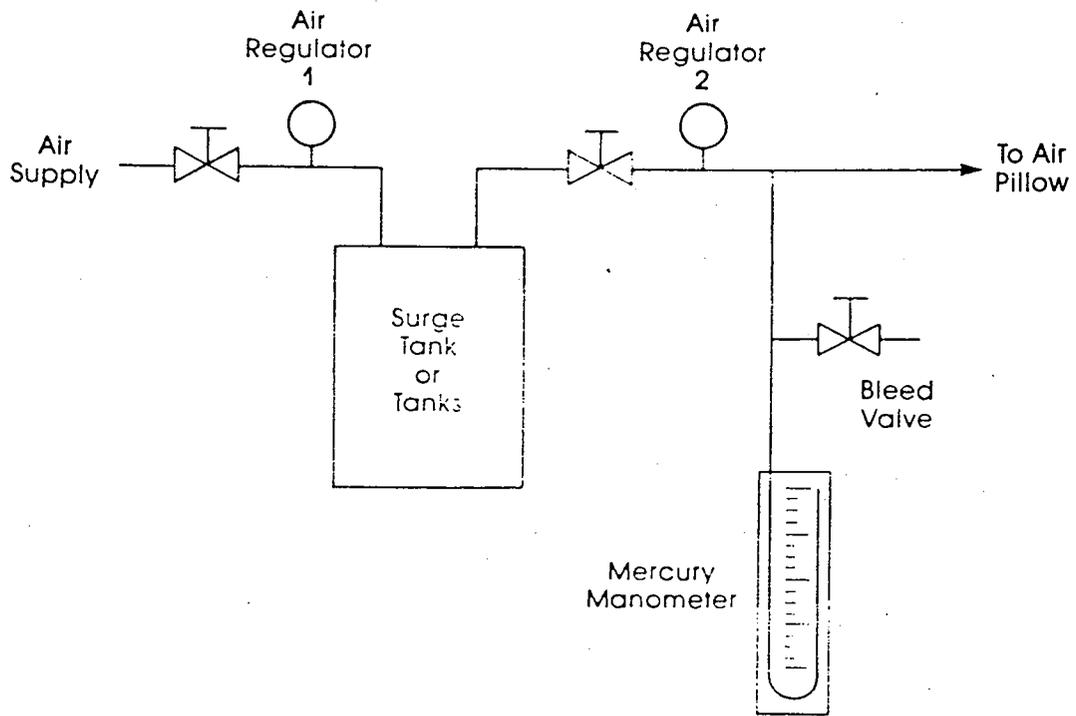


Figure 13. Air Supply to Pillow.

CRITERIA FOR FATIGUE DESIGN AS APPLICABLE
TO ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

Fatigue is a localized progressive behavior involving the initiation and propagation of cracks to final usually sudden fracture. In welded steel structures used for RBC's the initiation stage is non-existent because of the presence of small discontinuities introduced by all welding processes in or near the weldment - although good welding practice will minimize the number and size of these discontinuities.

The first part of this paper explains how existing codes handle design criterion for welded structures for fatigue, what is meant by allowable stresses, stress categories, stress range etc.

The second part will explain methods of improving fatigue life by controlling one or combinations of variables such as flaw size, residual stresses, stress concentration and will describe practical application of one such method. Two full size RBC shafts manufactured by above method has been successfully tested for 16 million cycles (equivalent to 20 years of life).

PART I

Research on fatigue has continued and is continuing steadily. In recent times much emphasis has been put on quantitative methods of handling crack propagation and final fracture. In this paper we shall first investigate the theory of fracture mechanics as applied to fatigue and review briefly the development of current provisions provided in codes.

PHENOMENON OF FATIGUE

The phenomenon of fatigue of steel can be divided into three distinct stages; initiation, steady growth, and rapid fracture of crack. (Figure 1) (1)

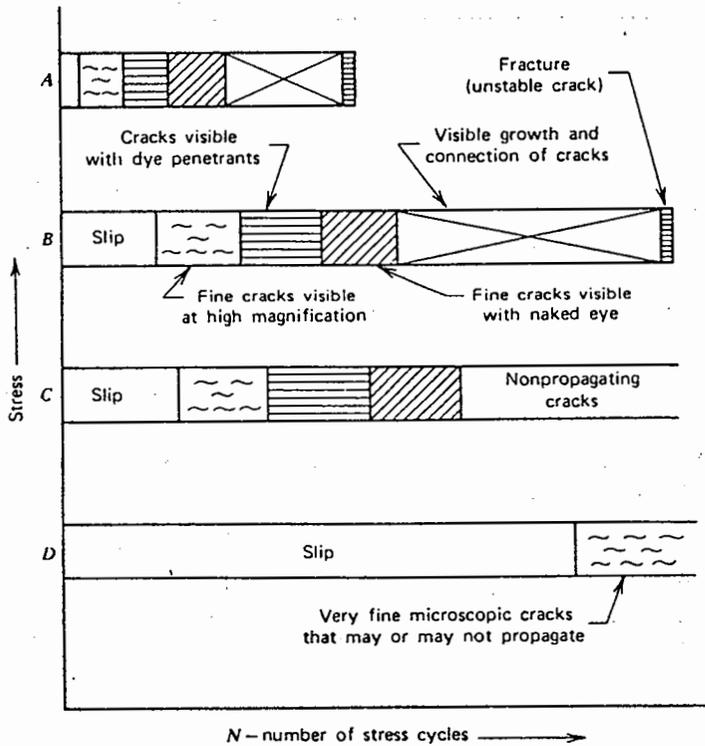
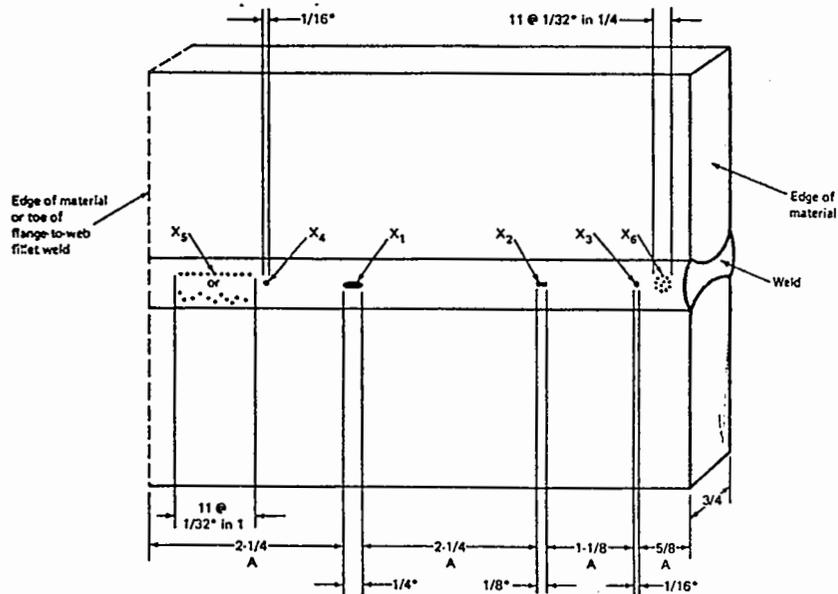


FIGURE 1. Schematic representation of the fatigue process.

In rolled beams, small flaws or microcracks start from slip planes of grain boundaries. For welded specimens the initiation stage is negligible because defects are always present in the weldment or at the weld metal-base metal interface, regardless of the quality control procedures and fabrication methods. As an example, discontinuity limits set by AWS for tension joints in bridges are shown in Figure 2. (2)



Notes:

1. A—minimum clearance allowed between edges of porosity or fusion-type discontinuities 1/16 in. (1.6mm) or larger. Larger of adjacent discontinuities governs.
2. X₁—largest permissible porosity or fusion-type discontinuity for 3/4 in. (19.0mm) joint thickness.
3. X₂, X₃, X₄—porosity or fusion-type discontinuity 1/16 in (1.6mm) or larger, but less than maximum permissible for 3/4 in. (19.0mm) joint thickness.
4. X₅, X₆—porosity or fusion-type discontinuity less than 1/16 in.

Interpretation:

1. Porosity of fusion-type discontinuity X is not acceptable because it is within the minimum clearance allowed between edges of such discontinuities.
2. Remainder of weld is acceptable.

*Defect size indicated is assumed to be its greatest dimension.

FIGURE 2. Example of discontinuity limits allowed by AWS.

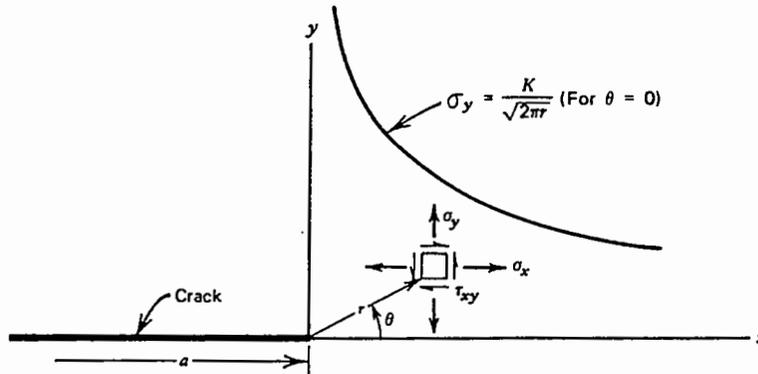
In fact it has been known that special efforts to completely eliminate these defects sometimes result in an inferior condition for fatigue. Also since welded joints constitute areas of high stress concentration these defects are unfortunately located at high concentration points caused by the geometry of the welded joints.

Present day design methods consist of limiting stress levels at which growth of the flaws already present in the welding is prevented or minimized. The resulting crack size developed during cyclic loading must be less than the size that will result in catastrophic failure.

The steady growth of cracks can be evaluated by using the theory of fracture mechanics. The steady state crack growth of the initial defects can be reasonably described by the crack growth rate equation:

$$\frac{da}{dN} = C(\Delta K)^n \text{ - - - - (1)}$$

Where a is the crack size, N the number of repeated application of stress, C and n material constants, and ΔK the range of stress intensity factor (Figure 3). Stress intensity factor should not be confused with the elastic stress concentration factor K_t , which is the ratio of the maximum stress at a notch to the nominal stress at the notch.



$$\sigma_y = \frac{K}{\sqrt{2\pi r}} \cos \frac{\theta}{2} \left[1 + \sin \frac{\theta}{2} \sin 3 \frac{\theta}{2} \right]$$

$$\sigma_x = \frac{K}{\sqrt{2\pi r}} \cos \frac{\theta}{2} \left[1 - \sin \frac{\theta}{2} \sin 3 \frac{\theta}{2} \right]$$

$$\tau_{xy} = \frac{K}{\sqrt{2\pi r}} \cos \frac{\theta}{2} \sin \frac{\theta}{2} \cos 3 \frac{\theta}{2}$$

$$\sigma_z = \tau_{xz} = \tau_{yz} = 0 \quad \text{for plane stress}$$

$$\sigma_z = \mu [\sigma_x + \sigma_y] \quad \left. \vphantom{\sigma_z} \right\} \text{for plane strain}$$

$$\tau_{xz} = \tau_{yz} = 0$$

Factor K is called stress intensity factor.

FIGURE 3. Elastic stresses near the crack tip ($r/a \ll 1$) (1)

The value of stress intensity factor depends on the loading, body configuration, crack shape and mode of crack displacement and can be expressed in a general form as:

$$\Delta K = F_e F_s F_w F_g S_r \sqrt{\pi a} \text{ --- (2)}$$

Where F_e is a crack front shape correction factor, F_s corrects for free surface condition. F_w corrects for finite width of a plate, F_g is the stress gradient correction factor which adjusts for the stress concentration caused by the detail geometry, and S_r is the stress range.

The constants C and n in crack growth equation (1) are characteristic of each material and are independent of the crack size in the steady crack growth stage. When equation (1) is integrated, it results in an expression for the fatigue life as:

$$N = \frac{1}{C} \int_{a_i}^{a_f} \frac{da}{(\Delta K)^n} \text{ --- (3)}$$

Where a_i and a_f are the initial and final size of the crack. Substituting equation (2) into equation (3) yields:

$$N = S_r^{-n} \left\{ \frac{1}{C} \int_{a_i}^{a_f} \frac{da}{[F_e F_s F_w F_g \sqrt{\pi a}]^n} \right\} \text{ --- (4)}$$

This expression indicates the stress range-fatigue life relationship.

We shall delve further on this equation and determine the variables that are critical for fatigue phenomenon. Left hand side of the equation denotes fatigue life in total number of cycles. The first term on the right hand side is stress range and the equation shows that the relationship between stress range and fatigue life has a slope of $-n$ if plotted on a log-log graph.

CONCEPT OF STRESS RANGE (4)

Stress range is the total range of stress a structural detail is subjected to. This means if the detail is subjected to 8000 psi tensile and 4000 psi compressive, the stress range on the detail is 12,000 psi. Although a microflaw would grow under a variable tensile stress only, compressive stresses are also included in the stress range for the following reason. All welding processes result in high tensile residual stresses. These stresses are at or near the yield point stress. A compressive stress in this area would only reduce the residual tensile stress but the actual stress would still be in the tensile zone. For example, if the residual tensile stress at some point near the weld is 24,000 psi and a 8,000 psi tensile stress and 4,000 psi compressive stress are subjected on the structure, the net stress at the point would be between 32,000 psi tensile to 20,000 psi tensile. This is the concept of stress range. It is a relatively recent concept and today used by all codes as the critical stress variable for all structural steel.

Getting back to the equation (4) we find variable n and c as material constants and are independent of the crack size in the steady crack growth stage. As long as the material is steel these factors will basically remain unaffected.

a_i is the length of initial flaw size. This is also very critical, (Figure 4 also shows effects of fracture toughness.)

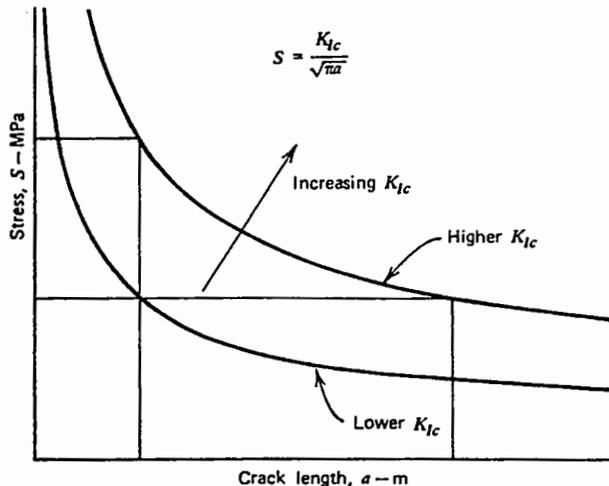


FIGURE 4. Influence of fracture toughness on allowable stress or crack size. (1)

perhaps the most unpredictable, undefinable, and inconsistent variable in the fatigue life prediction of welded structure. Normally the value of a_i for most weld toe flaws is less than .02 in. The quality and consistency is maintained by the procedure of production, obviously an automatic welding procedure would be more reliable than hand welding. We shall come back to this item in Part II.

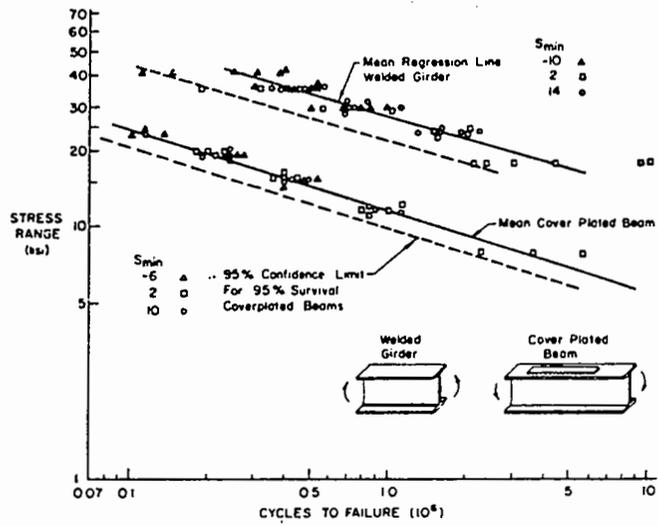
F_e, F_s, F_w and F_g have been defined before. Of all these numbers, F_g deserves a lot of attention. F_g is dependent on the stress gradient. The stress gradient is dependent upon the structural detail geometry—change in cross section or stress path control the gradient — the longer the change the sharper is the gradient.

To summarize, fatigue life of welded structure is primarily dependent on stress range, initial flaw size, stress gradient and final flaw size before fracture. To objectively find out the flaw size before designing a part is impossible and also there would be a scatter of flaw size present in the parts following any welding procedure. We now see the complexity of devising a design code.

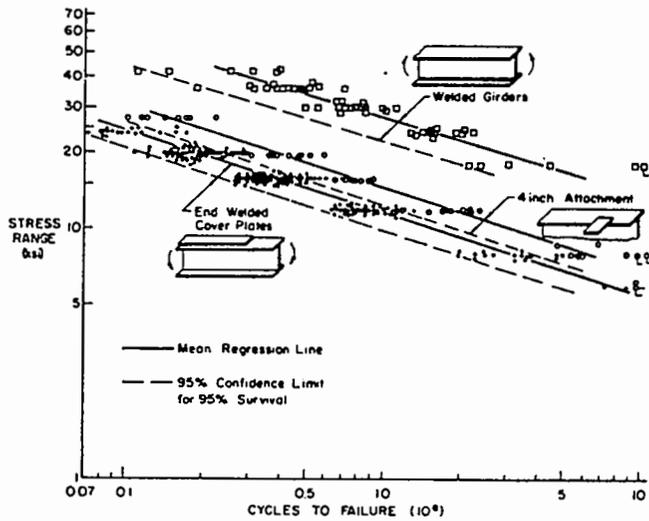
The present state of the art for design parameters for fatigue design takes an approach from experimental studies. In these studies a large number of test samples are made by following 'normal' welding procedure and tested at different stress ranges and the corresponding life cycles are noted. Three major areas that these tests are designed to address are:

1. Type of weld detail — takes care of stress intensity factor we discussed earlier.
2. Stress conditions — takes care of stress range we discussed earlier.
3. Number of tests — takes care of scatter of flaw size, residual stress and any other subjective variables of method of production.

Obviously we would expect a scatter in the results of these tests. The results of one particular detail is plotted on a log-log graph and the lower bound (95% confidence line) of the scatter points is used as the allowable stress ranges for life expectancy as required. Figure 5 shows such scatter points and 95% confidence line for 5 different details. (4)



Effect of minimum stress and stress range on the cycle life for the welded end of coverplated beams and plain welded beams.



Comparison of short welded attachments with coverplated and plain welded beams.

FIGURE 5.

The categories are termed as A, B, C, D, E and E' and correspond to plain material and rolled beams (A), plain welds and welded beams and plate girders (B), stiffeners and short attachments less than 2 in. length (C), long attachments - (4 in. length)- (D), and cover plated beams (E and E'). (Figure 6)

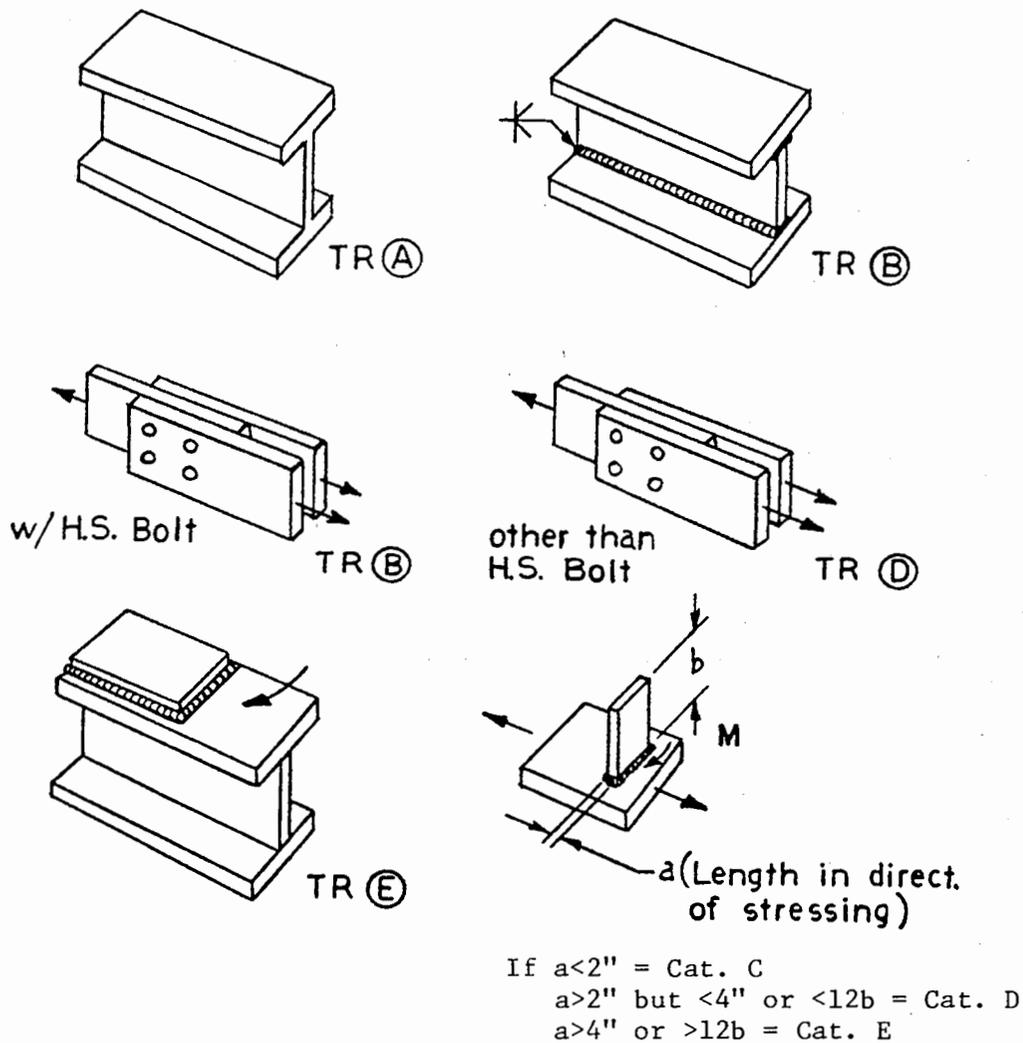


FIGURE 6. Some illustrations of categories used for fatigue. (5)

A typical curve is shown on Figure 7. The horizontal portion of allowable stress range - fatigue life curves are derived from an assumption that fatigue crack growth threshold exists. In small attachment detail it has been proved to exist by testing, although in some structural detail the horizontal portion of this curve may continue to slant slowly. (6)

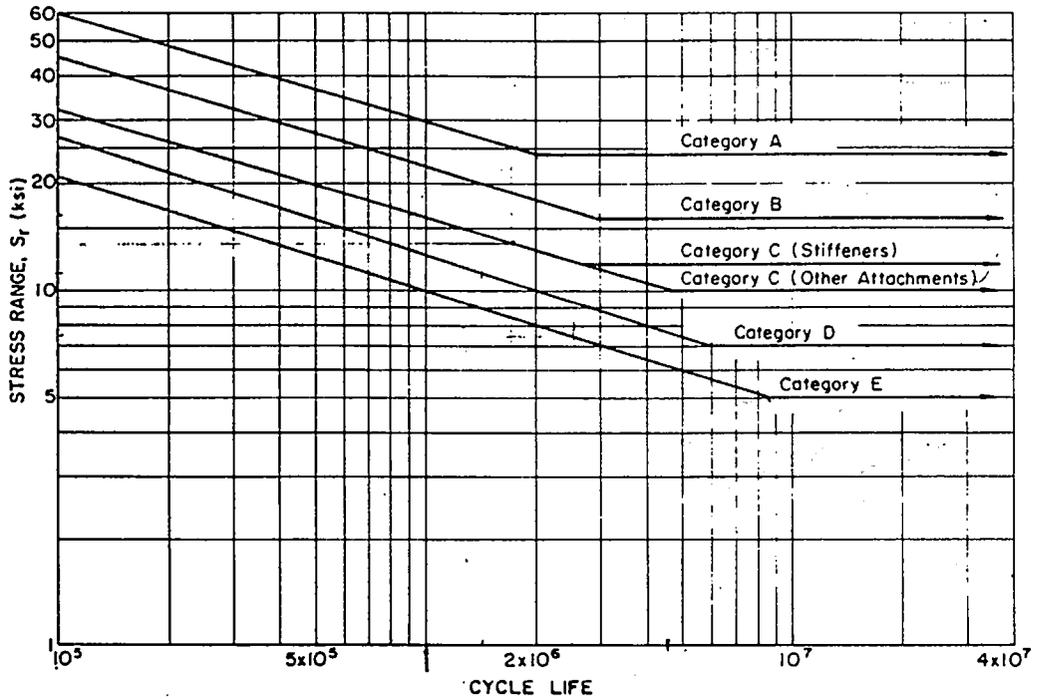


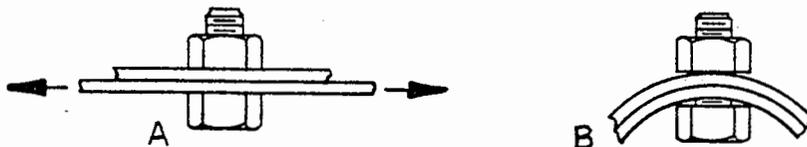
FIGURE 7. Design stress range curves for Categories A to E (4)

INTERPRETATION OF CODES

Some experience of fatigue behavior is necessary for interpretation of code data. We shall put forth some examples.

CASE I - A BOLTED JOINT

A bolted or riveted joint is normally considered category (B). We all know bridges built with rivetted joints have held very well over periods of years. One reason these rivetted joints held up so well is because when hot rivets cooled they shrank and introduced a clamping load around the holes. More recently high strength bolts introduce such clamp loads and a category (B) is allowable. But for a minute let us think about a hub on a circular pipe. Since the surface is not flat, tightening force by the bolt may not introduce a clamp force uniformly around the hole because of curvature. (Figure 8) A hole without clamp force is not category (B). It is a category (D). (4) Allowable stress for category (B) is 16,000 psi while for (D) it is 7,000 psi.



- A. Clamp load around bolt hole.
- B. Clamp load not easily obtainable.

FIGURE 8

CASE II

A beam with longitudinal attachment welded continuously along the length is a category (B). Consider a pipe with three bars welded along the length. In a bending mode (Figure 9) this is category (B).

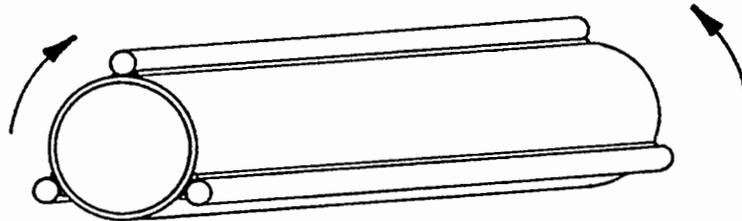


FIGURE 9. Bending mode loading of longitudinal stiffener.

For a moment let us think of the purpose of these bars. If these bars are used for holding loads in the circumferential direction, the loading is quite different. (Figure 10)

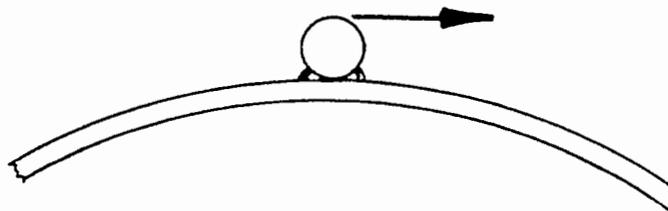


FIGURE 10. Loading on stiffeners in circumferential direction.

Here we have a load that is trying to put the weld in a tearing mode and the detail can be compared to cruciform joint. (Figure 11)

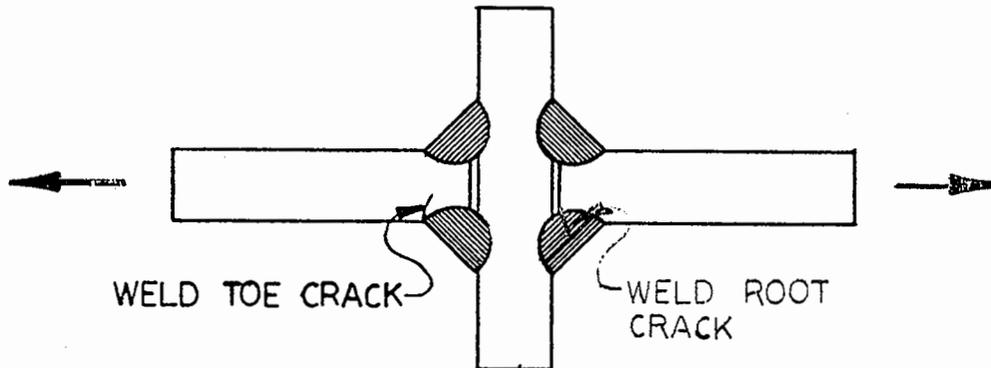


FIGURE 11. Typical Cruciform Weldment (7)

A fillet welded cruciform joint is category (F) and the allowable fatigue threshold stress range for category (F) can be as low as 7,000 psi on hot spot stress range for 15 million cycles, while for (B) it is 16,000 psi. Unfortunately in a cruciform detail, as the thickness of pipe is increased, the joint resistance to fatigue decreases.

CASE III

Some questions have been raised as to the correct category of transverse stiffeners. Consider a circular pipe with a transverse attachment. (Figure 12)

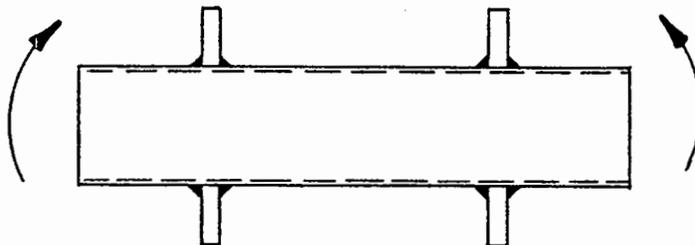


FIGURE 12.

Here the fillet is transverse to the stress path. This condition is certainly worse than A or B. Let us go back to equation (3) and examine the equation in relation to this attachment. We have noted before that one of the factors on which ΔK is dependent on is stress concentration at the weld. Stress concentration at any joint detail is dependent on the change of cross section. It is therefore important to take another look at the detail and consider the thickness of the attachment. Finite Element Method Studies have shown a strain distribution as follows: (Figure 13)

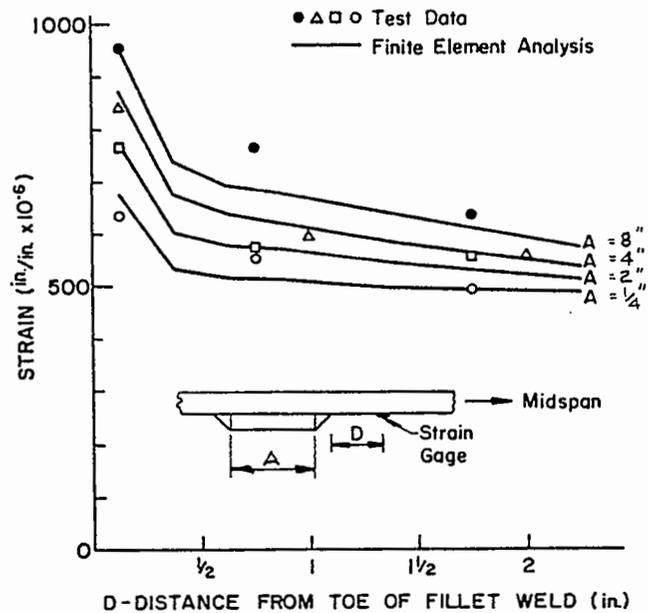


FIGURE 13. Strain distribution in beam flange near toe of transverse weldment of attachment. (6)

Obviously a 1/4" thickness will introduce significantly less stress concentration than a 2" thickness and therefore ΔK will be less, and a longer life is expected from the detail with a 1/4" thickness stiffener than with a 2" thickness. This is why AASHTO specifies that with less than 2" attachment the category to be used is C, and for 2" more the category is D. The allowable stress ranges for category C is 10,000 psi and for category D is 7,000 psi.

PART II

METHODS OF IMPROVING FATIGUE LIFE (8)

Since we have investigated the parameters for fatigue failure it would behoove us to consider ways of improving the fatigue-life. The relevant parameters are listed below:

1. flaw-size
2. residual stress
3. stress field

Extensive tests have been made to generate significant data on three methods generally applied to improve fatigue life.

GRINDING

Grinding the weld toe with a burr to provide a smooth transition and minimize the size of the initial discontinuities was the least reliable method. Some improvement was noted at the lower stress range levels, but none was observed at the highest level of stress range.

PEENING

In this method the weld toe is mechanically air-hammer peened with a blunt tool. Peening is continued until the weld becomes smooth. If a crack is visually apparent peening is continued until the crack is no longer visible.

Peening was observed to be most effective when the minimum stress was low. This was true for as-welded and precracked detail, and appeared to be directly related to the compressive residual stresses introduced by the peening process. When peening was carried out on unloaded beams, the application of a high minimum stress and/or high stress range decreased the effectiveness of the residual compressive stress that were introduced. Several tests were carried out on beams that were peened under a simulated dead-load condition. Under these conditions about the same improvement was noted at both high and low minimum stress levels as at higher stress range levels, thus indicating that to be effective, peening is to be done on a loaded structure.

GAS-TUNGSTEN ARC REMELT

The gas-tungsten arc remelt process involves remelting metal arc weld toe. The tungsten electrode is manually moved along the weld toe at a constant rate and melt a small volume of the fillet weld and base metal. Figure (14) shows a sketch of the gas tungsten arc remelting equipment.

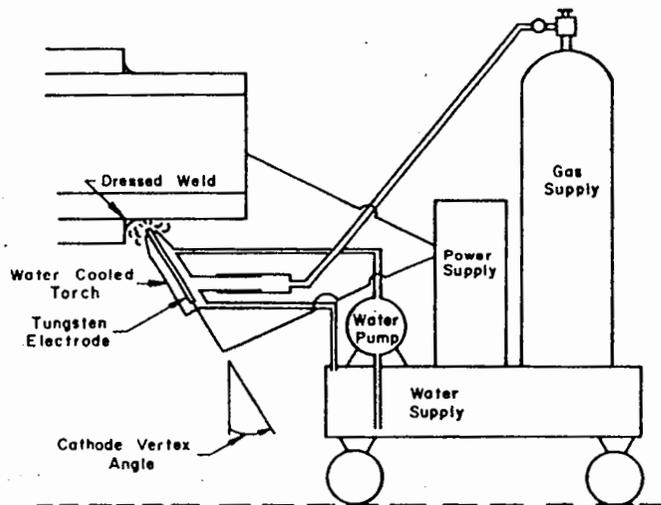
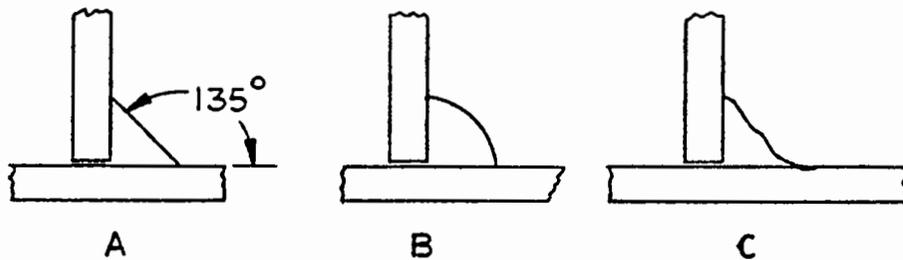


FIGURE 14. Sketch of welding equipment. (8)

This process achieves the following:

1. Removes non-metallic intrusion at the toe of weld by melting the metal and floating them up on the surface.
2. Removes micro-flaws at toe by melting the surrounding material.
3. Reduces stress concentration at toe by providing a smooth transition. (Figure 15)



- A. Depicts standard 45° fillet.
- B. Depicts weld profile not uncommon in production run. Notice sharp transition between weld profile and base metal.
- C. Profile after TIG remelt. Notice smoother transition between weld metal and base metal.

FIGURE 15.

The method is extremely useful in improving the fatigue-life of as-welded structures and can be used readily in the production of RBC shafts. Clow Corporation uses this method to improve fatigue life of wheel weldments.

The welding apparatus is a 300-amp d-c power source with drooping V-I characteristics. High frequency is used to start the arc. A Linde HW-18 water cooled torch with a .156 in. diameter, 2 percent thoriated tungsten electrode is used. This large diameter electrode ensures a sufficient depth of penetration to fuse all the cracks generated during initial welding. The shielding gas is 50-50 argon-helium mixture.

The wheels on RBC shafts are welded by automatic process in one station. The shaft assembly is then shot blasted to remove all mill scale and generally clean the area of all impurities. It is then put on a positioner and slowly turned while melting the toe with the tungsten electrode. The width of the melted zone is about 3/16" with one-half of it on weld metal and one-half on base metal. Impurities and gas pockets cleaned up by the molten puddle can be visually noticed by material floating on the puddle and by its occasional popping. The travel speed is approximately 10"/minute.

Toe remelt ensures no failure at weld toe and can reliably improve the fatigue strength one design category. Although for normal load RBC shaft design (50,000 lbs. maximum) such method of improvement is not required, Clow does this anyway. (8)

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THE AIR FORCE EXPERIENCE IN FIXED-FILM
BIOLOGICAL PROCESSES

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FOREWORD

Some 30 fixed-film biological wastewater treatment plants are located on Air Force installations. Most of the fixed-film process plants are trickling filter systems, except one with the contact aeration process and two with the rotating biological contactor(RBC) process.

This paper will discuss the Air Force experience in upgrading the contact aeration process plant with plastic media; the application of trickling filter to phenolic wastewater treatment; and the "troubleshooting" of an RBC wastewater treatment plant.

I. CONTACT AERATION PROCESS PLANT UPGRADING

INTRODUCTION

Upgrading of existing wastewater treatment facilities may be required in order to meet more stringent effluent discharge requirements or to handle higher loadings under existing effluent limitations. Depending upon the type of existing treatment facilities, the solutions to these problems may require process modification or significant expansion and/or modification of the existing facilities.

The wastewater treatment facilities at one Air Force installation required upgrading to improve effluent quality. This plant employs contact aeration, a two-stage process with contact surfaces submerged in the aeration tanks (see Fig. 1). The operation of the contact aeration process is quite simple, and does not require either sludge return, as in the activated sludge process, or effluent recirculation, as in the trickling filter process.

Because of this simplicity of operation, process modification alone in a contact aeration plant would not be an adequate upgrade. However, replacement of the original media with media of a high specific surface area will be very beneficial in improving treatment plant performance without incurring significant capital expenditure.

The plain asbestos plates originally installed at the treatment plant were replaced with a high specific surface area, module-type plastic media in April 1979. This paper will discuss the contact aeration process first and then use the performance data to evaluate the upgrading of this process with plastic media.

LITERATURE REVIEW OF THE CONTACT AERATION PROCESS

A. Brief History of the Contact Aeration Process

"Contact aerators" are contact beds that are continuously submerged in the wastewater they treat. Contact materials in the past included stone, coke, laths, movable pieces of cork or wood, corrugated

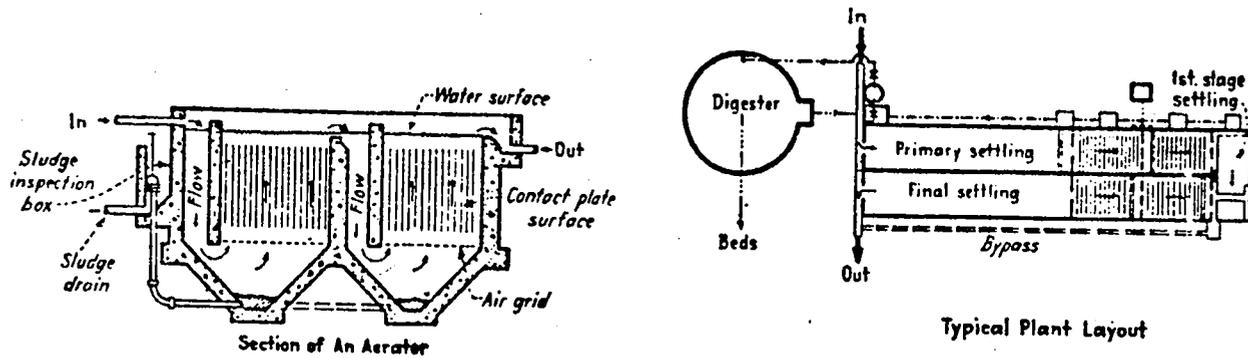
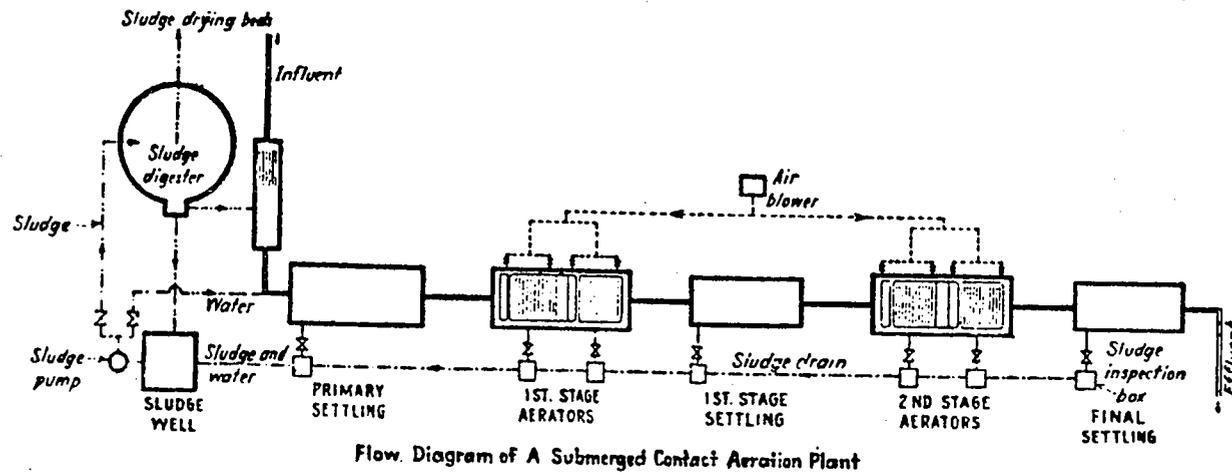


FIG. 1. CONTACT AERATION WASTEWATER TREATMENT PLANT

sheets of aluminum, artificial stone or ceramic materials of special shapes.

In 1929, Buswell and Pearson (1) suggested a "Nidus Tank" arrangement which was constructed to allow for contact surface treatment in two stages separated by intermediate sedimentation. The contact surface was provided by mats woven from veneer or basket strips and placed vertically in the aeration tank. Compressed air was introduced through perforated pipes placed underneath the "Nidus (Nest)" racks.

Between 1930 and 1938, Clyde C. Hays, City Chemist of Waco, Texas, developed a new flow diagram and patented this contact aeration process as the "Hays Process" (2). The first municipal contact aeration plant in the United States was constructed at Elgin, Texas, in 1939 (2). This plant utilized rock as the contact media.

During the next few years many improvements were effected. One of the most important of these was propounded by H. B. Schuehoff (3), who proposed the use of a series of flat asbestos panels in place of the rock media formerly used in the submerged filters. These contact surfaces were of $\frac{1}{4}$ in, 4 ft by 8 ft asbestos-cement plates placed on $1\frac{1}{2}$ in centers (3,4). The tops of the plates were submerged about 4 in. The aeration tank side water depth was about 9.5 ft with a cone-shaped bottom for the collection of sludge.

Over seventy Hays Process installations were in operation by 1943, including about fifty Army installations and a few Navy installations (2,3).

According to the Subcommittee of the Committee on Sanitary Engineering, National Research Council (NRC) (4), the contact aeration plants could obtain 80 to 95 percent BOD_5 removal under favorable conditions of loading.

However, when strong, stale sewage had to be treated, or where difficulties developed in the aeration system originally installed, effluents were unsatisfactory and odors became intense. Due to the difficulties of operation of contact aerators, high maintenance labor require-

ments, and more than occasional odor nuisances, NRC suggested that contact aerators were less desirable for use in military camps than were trickling filters. For this reason, the military contact aerators were gradually phased out and replaced by trickling filter or activated sludge treatment plants.

However, the evaluation of 27 contact aeration plants in New Jersey indicates that the contact aeration process was fundamentally sound (5). It was capable of efficient sewage treatment and, with correct design parameters and diligent operation, it could accomplish better than 90 percent removal of both suspended solids and BOD₅.

In 1967, a so-called "Fixed Activated Sludge Process" was studied (6). This system was actually a contact aeration system, but used plastic net panels as the contact surface. According to this study, this process could treat petrochemical wastes and soft drink bottling waste efficiently.

McCarty (7) used an upflow submerged filter to study nitrification in 1971. Although the flow pattern was upward in his study, the basic principle was similar to the contact aeration process.

A recent study in the application of contact aeration systems in biological nitrification (8) examined two adjacent contact aeration plants: one plant, constructed in 1965, had a 0.3 mgd capacity, and the other plant, constructed in 1973, had a 0.8 mgd design capacity. Portions (about 0.2 mgd) of the final effluent from the 0.8 mgd plant were pumped into the first aeration unit of the 0.3 mgd plant. The results indicate that a removal of up to 0.8 lb NH₃-N/1000 sq ft/day, or an effluent ammonia nitrogen concentration of 0.1 - 0.5 mg/l, could be achieved.

B. Design Criteria of the Contact Aeration Process

The design criteria of the contact aeration process are as follows according to Steel (3):

1. Settling Tanks

Settling Tank	Detention Time (hr)	Overflow Rate (GPD/sq ft)
Primary Settling	2	750 - 1,500
Intermediate Settling	1	1,500
Final Settling	1	1,500

2. Aeration Tanks

Aeration Tank	Detention Time (hr)	Air Req'd (cu ft/gal)	BOD ₅ Loading (16/1000 sq ft/Day)
1st Aeration Tank	1.2	} total } 1.5*	} average } 6.4
2nd Aeration Tank	1.2		

*The air distribution is normally 60% in the 1st aeration tank and 40% in the 2nd aeration tank.

According to NRC studies, the treatment efficiencies of five contact aeration plants at US Army posts had the following relationships:

$$E_s = 100 / \{ 1 + 0.225 [L / (At)]^{0.75} \} \dots\dots(1)$$

in which

E_s = percent reduction of BOD₅ based on settled sewage (%)

L = BOD₅ loading (lb BOD₅/day)

A = Contact surface area (1000 sq ft)

t = aeration time (hr)

The efficiency E_s in Eq (1) is based on primary settled sewage. To calculate the overall treatment efficiency, the primary settling efficiency should be included.

If the aeration time t = 2.4 hr, and the primary settling efficiency E_p = 35 percent, Eq. (1) can be rearranged as follows:

$$\frac{L}{A} = 15.56 \left[\frac{100-E}{E-35} \right]^{1.34} \dots\dots\dots(2)$$

in which

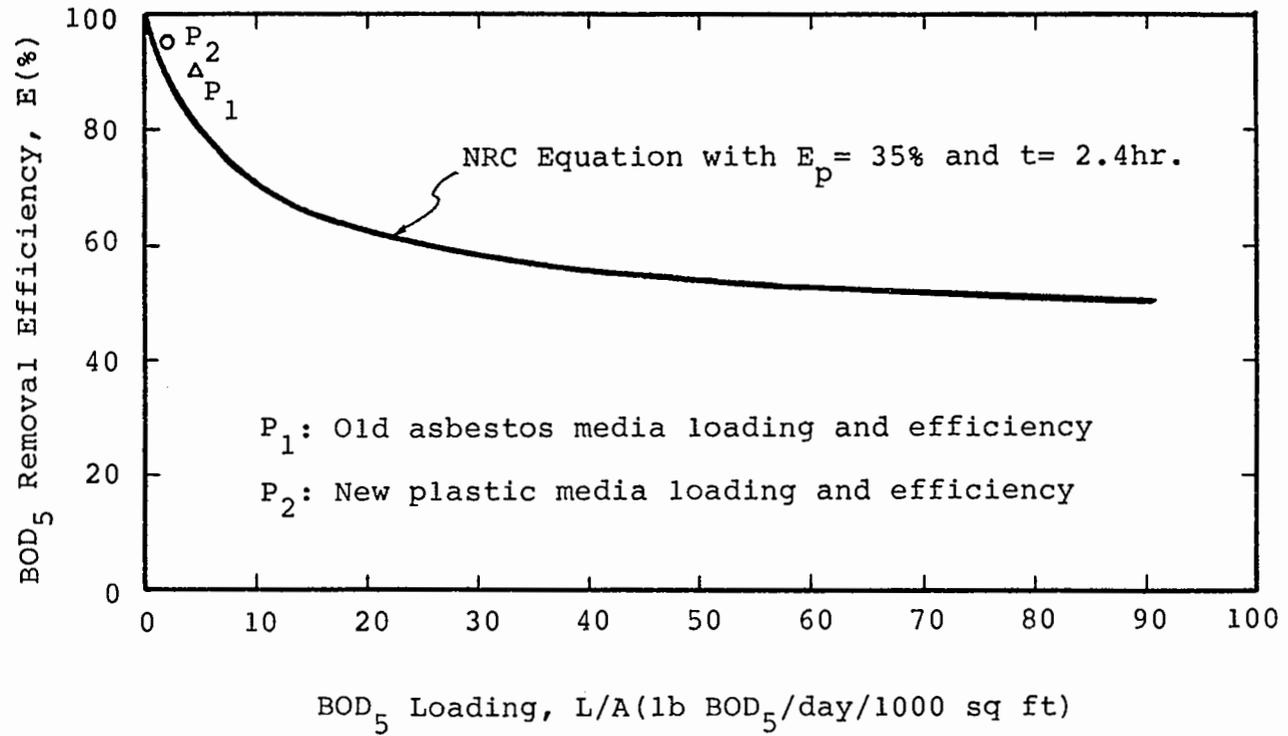
L = BOD₅ loading (lb BOD₅/day/1000 sq ft)

E = overall treatment efficiency (%)

From Eq (2), the BOD₅ loading versus overall treatment efficiency can be plotted as in Figure 2.

WASTEWATER TREATMENT PLANT DESCRIPTION

The treatment plant is a contact aeration plant consisting of a primary settling tank, a first-stage aeration tank, an intermediate settling tank, a secondary-stage aeration tank, a final settling tank, a chlorine contact tank, two polishing lagoons, and the irrigation

FIG. 2. BOD₅ LOADING VS. TREATMENT EFFICIENCY

lake. The flow diagram is shown in Figure 1 and the treatment unit descriptions are tabulated in Table I.

The original contact media in the aeration tanks consisted of about $\frac{1}{4}$ in thick, 4 ft by 8 ft plain asbestos plates hung vertically, with approximately $1\frac{1}{2}$ in spaces between the plates. The contact surface area was estimated to be 16 sq ft/cu ft. The new media used is of Koro-Z honeycomb[®] plastic made by the B. F. Goodrich Company. The specific surface area of this media is 44 sq ft/cu ft, with a void volume of 97 percent. The module dimensions are 2 ft x 2 ft x 4 ft.

The source of wastewater is domestic sewage generated by the base proper and family housing. The sewage flow rate during the surveying period varied from approximately 400,000 GPD to 450,000 GPD.

A dye test was performed at the two aeration tanks for a hydraulic distribution evaluation. The results indicated that the flow distribution through the honeycombed plastic media was quite even.

RESULTS AND DISCUSSIONS

The treatment efficiencies of the treatment plant with the asbestos plates (old media) and with the honeycomb plastic media (new media) are listed in Table II. Final effluent is the effluent from the secondary settling tank.

All of the analytical procedures were performed in accordance with the Standard Methods for the Examination of Water and Wastewater, 14th Ed (9).

TABLE I. DESCRIPTION OF THE CONTACT AERATION TREATMENT PLANT UNITS

Unit Parameters	Primary Settling Tank	First Aeration Tank	Intermediate Settling Tank	Secondary Aeration Tank	Final Settling Tank	Chlorination Tank
Dimensions (Wft x Lft x Hft)	12.25x67x8	12x33x8*	12x38x8	10x33x8*	10.25x69x8	7.4x10.25x8
Volume (gal)	49,100	23,700*	27,290	19,750*	42,320	4,540
Detention Time** (hr)	2.8	1.3	1.5	1.1	2.4	0.256 (15.4 min)
Surface Overflow Rate** (GPD/sq ft)	518	--	932	--	600	--
Contact Surface Area (sq ft)	--	101,380 (50,690)***	--	84,480 (42,240)***	--	--

*Contact media dimensions or volume

**Based on the flow rate of 425,000 GPD

***The original contact media before replacement

TABLE II. TREATMENT PLANT PERFORMANCE BEFORE AND AFTER
UPGRADING

Parameter (mg/l)	Raw Wastewater	Primary Effluent	Intermediate Effluent		Final Effluent	
			Old Media	New Media	Old Media	New Media
BOD ₅	186	121	--	40	18	9
COD	234	141	--	70	--	30
SS	181	110	--	33	14	6
NH ₃ -N	23.5	23.4	--	21.2	--	12.3
TKN	28.4	27.6	--	23.3	--	15.3
LAS	1.7	--	--	--	--	0.3

A. BOD₅ Removal

The BOD₅ removal efficiency was 90 percent when the original asbestos plates were in use. The BOD₅ loading was 4.62 lb BOD₅/day/1000 sq ft. The present BOD₅ removal efficiency is 95 percent after the original asbestos plates were replaced with honeycomb plastic media. The BOD₅ loading with this new media is 2.31 lb BOD₅/day/1000 sq ft.

The primary settling removal efficiency at this plant is 35 percent and the contact aeration detention time is 2.4 hr. If these two known values are incorporated into the NRC equation for the contact aeration process; Eq (1), the expected treatment efficiency at the sewage treatment plant would be as shown in Figure 2. However, the actual treatment efficiency at the treatment plant, as shown in Figure 2, with P₁ and P₂, is better than what the NRC equation predicted.

B. Suspended Solids (SS) Removal

The SS removal efficiency was improved from the original 92 percent up to 97 percent after media replacement. The present effluent SS concentration is 6 mg/l, which is excellent for a secondary treatment system.

C. Nitrification

Nitrification started to increase slowly two months after plastic media installation. The present ammonia nitrogen removal efficiency is up to 48 percent. Unfortunately, the nitrogen removal capability of the system before media replacement had not been analyzed. The only nitrogen data available are the analyses performed in November 1974, which indicated that there was no nitrification at all (10).

The major portion of the ammonia removal at this plant is in the secondary aeration tank. The overall ammonia nitrogen loading rate is 0.446 lb NH₃-N/day/1000 sq ft, and the removal rate is 0.077 lb NH₃-N/day/1000 sq ft in the first aeration tank, and 0.373 lb NH₃-N/day/1000 sq ft

in the second aeration tank. The conventional activated sludge system with an even longer aeration time, e.g., 4-8 hr, cannot achieve much nitrification. Therefore, this contact aeration system can out-perform conventional activated sludge, not only obtaining a very high BOD₅ removal, but also achieving a substantial degree of nitrification.

D. Dissolved Oxygen (DO)

The DO content of the final effluent, before the old air diffusers were replaced, was normally 0 mg/l, and occasionally 2-4 mg/l. The old, partially clogged, air diffusers were replaced with the Activator Hydro-Check Air Diffusers, Model 37. The final DO has increased to above 3-4 mg/l most of the time.

CONCLUSION

The treatment plant performed excellently after the old contact media and air diffusers were replaced. The upgraded plant can achieve not only 95 percent BOD₅ removal and 97 percent SS removal, but also about 50 percent nitrification and it provides a final effluent DO of 3-4 mg/l most of the time. A trickling filter system or an activated sludge system could achieve this kind of BOD₅ and SS removal only with good design and under carefully controlled situations. Nitrification, however, is hardly achievable in a secondary trickling filter system, or a secondary conventional activated sludge system, under comparable design criteria.

The contact aeration system is also easier to operate, because there is no need for recirculation or sludge return. There have been no "high maintenance labor requirements," or "more than occasional odor nuisances," as described in the NRC committee report (5), in this contact aeration system in past years, or in the 24 months after the new media were installed.

II. TRICKLING FILTER SYSTEM FOR PHENOLIC WASTE TREATMENT

INTRODUCTION

The Air Force maintains five Air Logistic Centers which are responsible for the maintenance, repainting and repainting of operational aircraft and ground support equipment. Aircraft and ground equipment are stripped (depainted) periodically to prevent corrosion of aircraft surfaces. A viscous paint remover/stripper is brushed or sprayed on the surface and is left there for a period of time while it swells, wrinkles, and softens the paint, thus lifting it from the surface. The paint remover and paint particles are then rinsed from the aircraft with a high-pressure water stream. This constitutes the source of the wastewater.

The type of paint system (topcoat and primer) on an aircraft dictates the type of paint remover required. The Air Force currently employs polyurethane topcoats with epoxy primers, which require paint removers containing significant concentrations of phenols (see Table III) (11). The concentration of the contaminants of wastewater will vary depending on the phenolic paint stripper and the amount of rinse water used. Working from the characteristics of the wastewater (phenol concentration) and the concentration of phenols in the paint stripper, it was estimated that each gallon of paint stripper is rinsed with between 45 to 75 gallons of water (12). The depainting of a B-52, strategic bomber, for example, requires approximately 3,350 gallons of paint stripper. If 60 gallons of rinse water are required per gallon of paint stripper, approximately 200,000 gallons of wastewater are generated from each B-52 depainting operation.

The wastewater generated from the depainting of aircraft and ground equipment represents the only major source of phenolic waste within the Air Force. Disposal of phenolic paint stripping waste is a serious and ever increasing problem facing the Air Force. Therefore, the Air Force is continuing to evaluate alternative processes in order to meet current and future discharge standards in a cost-effective manner.

TABLE III

ANALYSIS OF PAINT STRIPPER AND PAINT STRIPPING WASTEWATER

NOTE: All values in mg/l (except pH)

COMPONENT	PAINT STRIPPER	PAINT STRIPPING WASTEWATER
PHENOL	200,000	1040 - 4060
METHYLENE CHLORIDE	600,00	75 - 2000
SURFACTANTS	100,000	120 - 4000
PARAFFIN WAX	50,000	- - - - -
METHYL CELLULOSE	40,000	- - - - -
WATER	10,000	- - - - -
CHROMIUM TOTAL	2,400	17.5 - 59.5
CHROMIUM ⁺⁶	2,400	- - - - -
TOTAL PHOSPHATE (AS P)	- - -	10 - 28
SUSPENDED SOLIDS	- - -	107 - 303
VOLATILE SOLIDS	- - -	458 - 2700
TOTAL SOLIDS	- - -	800 - 3830
COD	- - -	9200 - 36400
COD FILTERED (0.45u)	- - -	7250 - 35100
TOC	- - -	2710 - 14400
TOC FILTERED (0.45u)	- - -	2520 - 13600
OIL AND GREASE	- - -	8.4 - 66.3
pH	- - -	8.0 - 8.5

The Air Force studies demonstrated control technologies including ozone and permanganate oxidation (12), activated carbon absorption (11, 12), and biological treatment (13, 14, 15). Blum (16) evaluated activated carbon absorption, several chemical oxidation processes, and biological processes for phenolic wastewater treatment. The results indicate that biological processes are by far the most cost-effective alternatives. Therefore, the aerated trickling filter, which will be described later, and the rotating biological contactor system are recommended for aircraft paint-stripping wastewater (16).

LITERATURE REVIEW OF BIOLOGICAL TREATMENT OF PHENOLIC WASTE

Many aerobic bacteria and fungi are capable of using aromatic compounds as the sole source of carbon and energy. Therefore, phenol removal can be achieved by biooxidation processes.

However, at high concentrations, phenol is toxic to most microorganisms. Some phenol biodegradation studies reported substrate inhibition at phenol concentrations above 100 mg/l (17, 18, 19) while others indicated that no substrate inhibition was evidenced for phenol concentrations up to 360 mg/l in one study (20) and up to 1000 mg/l in another study (21).

Biological treatment of phenolic wastes has been applied in petrochemical plants and has proven to be economical and reliable (22, 23, 24). One activated sludge pilot plant for treating weak ammonia liquor from a coke plant could achieve 99.9 percent removal at an influent phenol concentration of 3,500 mg/l (25). A full-scale activated sludge plant was reported to treat a chemical plant's waste from an influent phenol concentration of 1,026 mg/l down to an effluent phenol concentration of 0.35 mg/l (25). Phenolic compounds in coal gasification wastewater can also be removed more than 99 percent by the activated sludge system (26, 27, 28).

Some oil companies have employed trickling filters to remove phenol, and several such plants were reported to remove 88% to 98% of the phenol (23). Cooling towers, which are similar in operation to trickling filters, have also been employed to reduce the phenol concentration in

wastewater. Efficiencies of phenol removal both in forced and induced draft cooling towers were reported to be from 99.4 to 99.9% for phenolic concentrations ranging from 10-70 mg/l (20). From laboratory and pilot plant studies, Reid, et al. (30) found that phenols can be treated successfully by biological slimes in concentrations as high as 7,500 mg/l.

A tapered fluidized-bed bioreactor system, which uses coal or sand of about 30/60 mesh particle size, has been tested for coal conversion wastewater treatment (31, 32). The phenol removal efficiency of this type of reactor is reported to be 99.5%.

THE AIR FORCE STUDIES IN TRICKLING FILTER PROCESS

As mentioned previously, fixed-film biological processes are recommended for aircraft paint stripping wastewater in Air Force installations (16). The trickling filter process, however, is by far the most popular phenolic wastewater treatment system employed in the Air Force Logistic Centers.

For example, one Logistic Center uses a 65 ft diameter, 23 ft high plastic media trickling filter. The average influent phenol concentration is 22.0 mg/l and effluent phenol concentration is 0.70 mg/l, which represents 96.8% removal efficiency. To meet a more stringent discharge requirement, a granular activated-carbon process is being designed as the final polishing system to improve the removal efficiency.

The most recent Air Force study on paint stripping waste treatment is an "aerated" trickling filter (see Fig. 3), which utilizes forced aeration and plastic bio-ring support media (14). The system is able to reduce phenol concentrations from 1000 mg/l to less than 1 mg/l. The advantages of this system are that the growth media is not seriously affected by shock loading or routine 12-hour and 72-hour overnight and weekend "down" periods, and that the paint stripping wastewater could be continually recycled over the media until the desired effluent quality is attained.

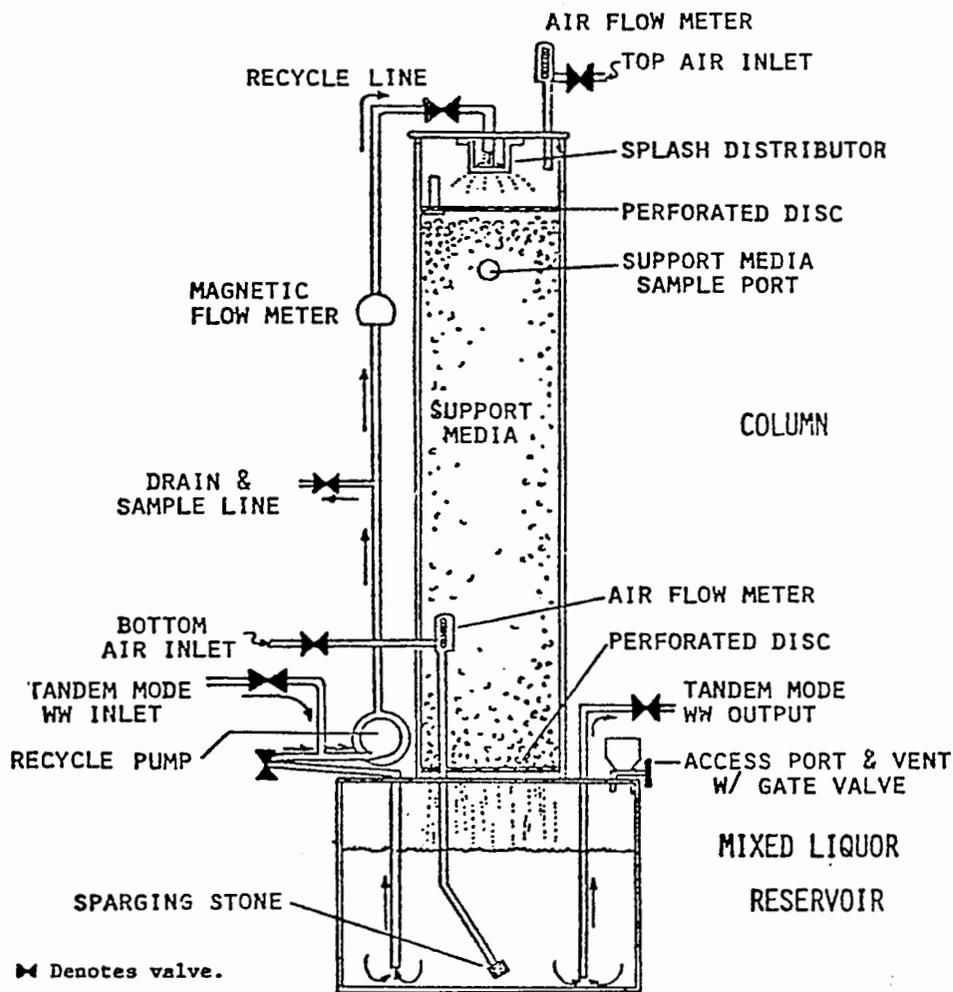


FIG. 3. AERATED TRICKLING FILTER

A 4th diameter, 6 ft high pilot scale "aerated" trickling filter study is currently being planned. This study will generate the design criteria and scale-up factors for a full scale treatment plant design.

III. THE "TROUBLESHOOTING" OF AN RBC TREATMENT PLANT

INTRODUCTION

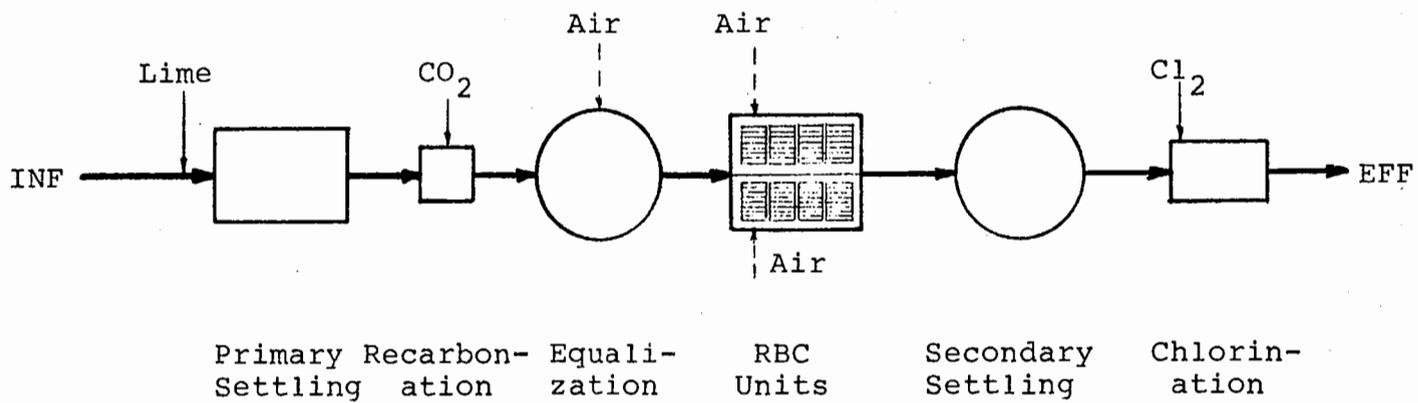
A rotating biological contactor (RBC) wastewater treatment plant in one Air Force installation had difficulty meeting the expected performance, so a "troubleshooting" study began.

The RBC plant consists of primary settling, with lime added for phosphate removal; recarbonation; equalization; the RBC; secondary settling; and chlorination. The schematic flow diagram is shown in Figure 4. The RBC system has two parallel series. Both series have four stages, each with an available contact surface of 60,000 sq. ft. and with baffle wall between each of the stages.

The operation of the RBC plant before the "troubleshooting" was described as follows:

Lime slurry was added at the primary settling tank inlet to raise the pH to 9.2 and to form a phosphate precipitate which settled out in the primary clarifiers. The primary effluent was discharged to a recarbonation tank to reduce the pH to 8.5. The recarbonated effluent was discharged to an equalization basin and then to the RBC system. The effluent from the RBC system flowed to the old secondary clarifiers for biological solid separation and then to the chlorination chamber before its discharge to the receiving waters.

Table IV shows the performance of the RBC wastewater treatment plant before the "troubleshooting".



Legend:

- Before and during study period
- - -→ Added during study period

FIG. 4. SCHEMATIC FLOW DIAGRAM OF THE RBC WASTEWATER TREATMENT PLANT

TABLE IV. THE RBC PLANT PERFORMANCE BEFORE "TROUBLESHOOTING"

PARAMETER	Raw Wastewater	Primary Effluent		Secondary Effluent	Total Removal
	mg/l	mg/l	% Removal	mg/l	%
Total BOD ₅	171	129	25	28	84
Soluble BOD ₅	85	83	2	18	79
Total COD	294	239	19	65	78
Suspended Solids	200	67	67	27	87
Total Phosphate (as P)	6.3	4.9	22	4.5	29

Note: Ave. flow rate = 0.82 MGD

PROBLEM INVESTIGATION

The "troubleshooting" study began in June 1980 with an intensive field investigation and sampling which revealed several problem areas relating to the RBC system:

A. A High Soluble BOD₅/total BOD₅ ratio in wastewater--

The soluble BOD₅/total BOD₅ ratio in wastewater is approximately 50%, while the typical municipal wastewater is about 22 percent (34). The implications of this high soluble BOD₅/total BOD₅ ratio are two fold: (a) the BOD₅ removal in primary settling tank is low, even with lime addition and (b) the RBC system design was based on soluble BOD₅ loadings (35); therefore, high soluble BOD₅ in wastewater would increase contact surface area requirement.

B. The carbon dioxide (CO₂) transfer efficient was low in the recarbonation tank --

The CO₂ transfer in the recarbonation stage was only about 50 percent efficient due to the improper diffuser selection. Therefore, the pH of recarbonated wastewater was as high as 8.5 rather than the expected pH range of 7.5 to 8.0.

C. The Equalization tank was not aerated --

Use of an unaerated equalization tank will not only allow solids to settle out, but will also turn the wastewater septic. The septic wastewater will adversely affect the performance of the subsequent biological treatment process, here, an RBC system.

D. An organic overloaded RBC system with a short-circuiting hydraulic pattern in the RBC holding basins --

The RBC system design is based on soluble BOD₅ loading, according to the manufacturer's design manual (35). The design criterion for the first stage of RBC system is 4.0 lb soluble BOD₅/day/1000 sq. ft., while the plant's loading

is approximately 4.70 lb soluble BOD₅/day/1000 sq. ft. for the first stage. Therefore, the RBC system is overloaded at the present loading condition.

The baffle wall between each stage of RBC unit had a 6-in opening along the bottom of the baffle wall. The dye test indicated that there was a serious hydraulic short-circuiting due to this improper design.

E. Hydraulic overloaded secondary clarifiers --

The overflow rate of the old secondary clarifiers was approximately 700 gpd/sq. ft., which exceeded the RBC manufacturer's recommended 500 gpd/sq. ft. overflow rate for meeting the 10 mg/l suspended solids effluent requirement (35).

CORRECTIONAL MEASURES

The following corrective measures have been implemented or are being evaluated:

A. The re-design of CO₂ diffuser system in the recarbonation tank --

The coarse bubble CO₂ diffusers were replaced with finer bubble diffusers. The pH of wastewater after the CO₂ diffuser replacement is maintained at 7.8-7.9 range. The adjustment of the pH to this value will permit development of a more desirable community of microorganisms in the RBC system.

B. Aeration of equalization tank --

Aeration in the equalization tank has been implemented. Aeration of the equalization tank will not only keep the solids suspended, but will also keep the wastewater from becoming septic.

C. RBC system modifications --

Several measures have been implemented or tested in order to correct the problems associated with the RBC system.

The 6-in opening at the baffle wall has been reduced to 1½-in to alleviate the serious hydraulic short-circuiting situation.

Diffused air has been applied at the first stage of the RBC units to increase the oxygen supply. The optimum diffused air rate is still under study. The rotational direction of the RBC units has also been reversed from clockwise (following the flow direction) to counter-clockwise (against the flow direction) so that more mixing and more intimate contact between biological slime and wastewater can be expected.

D. Secondary clarifiers --

The old, hydraulic overloaded secondary clarifiers were replaced with two new, larger clarifiers, which provide much better settling of suspended solids.

TREATMENT PLANT PERFORMANCE WITH CORRECTIONAL MEASURES

The RBC plant performance improved substantially after the above-mentioned correctional measures were implemented. The secondary effluent quality before and after this "troubleshooting" is shown in Table V.

Although the RBC plant performance is improving, the final effluent quality is still not satisfactory, especially for phosphate and ammonia removal. A better BOD₅ efficiency will be provided in the near future in order to meet a more stringent effluent limitation. All of these requirements are presently under study.

TABLE V. THE RBC PLANT PERFORMANCE BEFORE AND AFTER "TROUBLESHOOTING"

PARAMETER	RAW WASTEWATER	SECONDARY EFFLUENT			
		Before		After	
	mg/l	mg/l	%Removal	mg/l	%Removal
Total BOD ₅	171	28	84	18	89
Soluble BOD ₅	85	18	79	12	86
Suspended Solids	200	27	87	7	97
Total P (as P)	6.3	4.5	29	3.4	46
NH ₃ -N	19.7	-	-	10.4	47

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Workshop On Research Needs for Fixed-Film
Biological Wastewater Treatment

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Dr. A.F. Gaudy, Jr.

I want to introduce the fellows here on the podium. My name is Tony Gaudy and I'm now with the University of Delaware. I spent a long time out at Oklahoma State University before coming to Delaware, and my job is to introduce the guys here, you know. And sitting next to me is a very dear friend of mine, I'm very happy to say a former student of mine who's become a very, very famous guy. He put Purdue on the map and now he's going to do the same for Clemson, Leslie Grady. Everybody knows him. If I gave you his academic leads, I'd have to give you the academic leads for the rest of them and I'm not too sure I know about it. As I know, Les has worked with Art Bush for his M.S., and then Art said, Tony, I'm going to put you on the map. And I'm going to send you a real good student, and Les came on up and did his Doctor's with me at Oklahoma State and you know Les's story ever since then. Then sitting next to him is a guy who's going to talk about the Submerged Biological Filter, Dr. Friedman of Syracuse University. I'm not sure about all of Art's lineage. I'm not even sure about his parentage but I do know that he did work with Ed Schroeder of the University of California at Davis.

Art is going to introduce the subject of the Submerged biological filter which is a process which has been around for quite some time, but was really brought to the fore in recent years by Professor Perry McCarthy and his group out at Stanford. It really blossomed that process.

And next to his is Ed Opatken whom I know works for the EPA here in Cincinnati and he's really a chemical engineer, but is one who's devoted himself mainly to the environmental field for many, many years and he got his degree from the University of Pittsburgh. Everybody knows of Ed's activities and Ed's going to make some comments and again introduce the subject of Rotating Biological Contactors which have become a

very important process for biological treatment.

And next to him is a guy who, I don't think I don't know, this kid hasn't been around too long. I've been very privileged to know this guy for pretty near 30 years. When I came into this field as a Civil Engineer, trying to get into sanitary engineering, Wesley Eckenfelder was well known in sanitary engineering. I suppose he must be about the same age as I am. But anyway, he's been around in the field a lot longer than me, you know. Of course everybody knows he's the guy who made Manhattan College change the name from Manhattan College to Manhattan University. Everybody remembers Wes from his days with Roy Weston and hooray for those days at Manhattan College and hooray for those days at the University of Texas and now for those days at Vanderbilt University where they have a tremendous program of research and education in environmental engineering. And Wes--when I was a graduate student and Wes was out there knocking them dead--he's been knocking 'em dead for a number of years now, you know. But he was the guy who brought the chemical engineering approach to environmental engineering, by golly. He was the only guy that had an equation that seemed to work. And God bless him, he's been working that same equation for 30 years. If there's any guy, not only in this country, but in this world, who could talk about the subject of trickling Infiltration, it's got to be Wes Eckenfelder.

This is our group today and it's quite a distinguished bunch. And before I turn it over to these fellows to make some introductory comments, before they basically turn it over to you to really carry this workshop forward. I want to explain a little about you fellows, you're probably saying, what the hell's Gaudy doing, introducing this bunch of guys here, and after all what did he ever do for Fixed Film Reactor. He's been an activated sludge man all his life. Well, I'll tell you, if it wasn't for Dow Chemical, I may have turned out to be a Fixed Film man myself. It's all Ed

Bryan's fault and Tom Power's fault that I'm not a Fixed Film man, you see. I don't know how many guys remember Ralph Liason. Ralph Liason was a Chairman of Civil Engineering Department when I was a graduate student at MIT. Ross McKinny was a young Prof there. Doc Sawyer was at the height of his powers as a Prof there and Professor William Stanley was the designer and that was the four man crew at MIT that have spawned an awful lot of people who are in education in our field. And I was sitting in Liason's class one day and as the students in that class wanted to do, I was doodling. Professor Liason was talking about trickling filters you know, and I got to thinking, why use rock? Let's use something--maybe a guy could take a piece of corrugated asbestos and you could bolt these things together and you could stack 'em and maybe you'd have a trickling filter. And I thought that might be a great idea. So I took it into Ross McKinny's first graduate students back in his days at MIT. So I took it into Ross and he said, "Say that's not a bad idea. It might work." So we took it down to Sawyer, and Sawyer said, "That's not a bad idea. . . . it might work." So I was finishing up my Master's and he said, "Maybe you ought to stick around a little bit and work on it, and we'll see what we can do with it." So that was 1953 and I tottled off to the Purdue. I forget what Purdue Industrial Waste Conference it was, but they tottled off to the '53 Purdue Industrial Waste Conference. In those days the Purdue Industrial Waste Conference, was still is, a damn good conference with a hell of a lot of drinking. They straggled back about a week, a week and a half later and they saw me and they had this long, long face, And I said what's the matter and they said, "There's something we have to tell you. There was a couple of guys at Purdue who gave a paper from Dow Chemical and they were talking about trickling filter media and it's not made out of asbestos, it's made out of plastic, and it wiggles like this." They said, "It wipes your idea out." So if it wasn't for Ed and Tom, I might make claim to be a Fixed Film man. But I can't.

For biological treatment, I imagine everyone in this room and most people around the country has come to realize that the only way we're going to do it is by understanding and by treating organisms to do the job. And it makes very little difference with we do in on a fixed field or we do it on a fluidized bed. The principles may not be quite the same. The biological principles are probably the same. The ecological principles may be different. And the engineering principles are going to differ because we have the problem of sticking it to a surface, adhering it to a surface, and the problem of getting it off the surface. And the reasoning involved of chemical, biological, and physical aspects but at the chemical core of all biological treatment processes they are similar and so people, presumably many of the people here, are devoting their efforts to understanding and applying biological treatment using a fixed biota whereas others may apply it using a fluidized biota but, nevertheless, it is a biota. And the basic bio-chemical principles that govern it are the same, so that there's really not all that much difference. That's my excuse for being here at any way.

Before I turn this over to these gentlemen, I want to recognize all the people who are here, you know, and I feel--well I've been privileged to be in it for over a quarter of a century now--and it's the kind of field that is populated by people who periodically do an awful lot of self-examination. We probably do more applied research in our field than any of the applied areas that I can think of. I don't know whether this is typical of engineers or not. Sometimes we do it almost to our detriment. There are times I think we'd be better off leaving well enough alone. But we persist in investigating. We persist in this process of self-criticism and it's good. It's the kind of thing--well it reminds me of a story that happened back during the French Revolution. And you know, back in those days, when the Revolutionaries figured someone did wrong they'd take him out and they'd have him a mock trial and they'd run him through the guillotine and as soon as there was a man of

the cloth said, "Well I really haven't done anything wrong. And I'd really rather be looking at heaven when I got it." So he goes on, lays on the block, and the drums roll, and down comes the blade and stops about a half an inch from his neck. And the crowd yells, "He beat the guillotine. Let him free." And they untied him and they let him up, you know. Well the other fellow who was a lawyer said, "by Golly, I want to take a chance on that. That was very good, I might get away with the ing." So they put him face up, the drums roll, down comes the blade, vomp--stops a half an inch from his adams apple and they let him free. And the engineer said, I don't know maybe I ought to try the same darn thing you know. And they laid him on his back, and he said, "Hey, I know why that thing doesn't work, there's a knot in the cord."

Tonight, I don't know if we're going to hang ourselves or cut our heads off, but by golly, we're going to look into some questions about fixed bed reactors. And, I don't know, who wants to lead off now. We got RBC's, we got trickling filters, and we got submerged fixed reactors and there's a number of people, there should be a number of people, if I read off the list I have here of the people we expected to be here, who have done work on those three reactors, it'd be quite a long list. The most names that I have are people who have done recent research on RBC's and if we go with what looks like it was of the most interest, at least in our preassessment of the workshop, let us start with Ed Opatken. Ed, would you like to say a few words about RBC's?

Mr. Ed Opatken

Join in any time you want. I guess the first RBC plant that construction grants build was in Gladstone, Michigan around 1972. Everybody see that plant? Bob did. Anybody else? Anybody see pictures of it? Anybody think they could tell the difference between that plant built in 1972 and, say, one built in 1981? Does anybody think

there's a difference?
Cover is different. Very good. Let's continue.
We've got covers here now. Is there anything
else?

The media.
The media is different. Stronger shafts and
bearings. O.K. Anything else?
Air supply.

So if we go back to the first one we do
find that we have made some advancement. Some
of the things we brought up which we've studied
or we haven't--should we decide the BOD(?)
should be used stage BOD?

Any ideas along those lines? Possibly.

I sometimes shiver when I see kinetic equa-
tions mixed with mass transfer equations. Back
in my days, there's only one limiting factor,
there's only one always. And the combination
of two is really not, in my opinion, of any
significance. Only one controls the system.
Equipment failure, have you seen any of those on
RBC's? What reason do you have for them? Any-
body have any ideas on shaft and media failure.
Have you had any papers on failures? Or in-
vestigations of them?

Some of the other aspects is, I think we
said, one of the parts we have is a facility
built in 1982 and we want to have a nice
plant available. The aesthetics are right
there. Do we have any controls at all on an
RBC? Suppose we tell an operator you know,
'operate this plant." If we get a high organic
loading, what are you supposed to do? Can you
do anything? Have we given them anything?

We can sit back and say we've gone a long
way. I've gone to one plant and have an opera-
tor say to me, "I've got four parallel trains
here and I'm sure that the two end ones are
getting more flow than the two little ones: and
all I can do is just look at it. I can't mea-
sure it. And I don't know what we should do to

improve that. We talk about DO (?). I think we hear the DO sags. We provide a measurement of it, and if we do, we hear the same excuse that it wouldn't be any good anyway because the operator wouldn't maintain it. Do we provide any flexibility on an RBC system besides I heard the supplemental air being added. Is there anything else that we can do to an RBC system that would provide an operator with the capability of altering it or do we design it into a new system so that if something occurred we could automaticall respond? I think there's a lot of room that we could work in. Do any of you have any suggestions where the effort should be concentrated the next five to ten years?

Mr. Frank Viteck: Well, I think there should be much more said about this idea of rotational speed. The manufacturers right now use the fixed rotational speed and are happy with it. And I would suggest we could go at higher RPM's and achieve certain treatment efficiency, but structurally, no one is willing to stick their neck out and say they can withstand that kind of rotation speed over a period of time and I think that that question isn't being resolved because the manufacturers aren't addressing themselves to it.

Mr. Tom Shore: Just the idea of rotational speed. Right now we're sitting at a happy rotational speed of 1.6 RPM and feel we can operate at that speed, without structural failure. But there isn't much interest at operating at higher rotational speeds and the pilot work has been done. It's old. It suggests treatment efficiencies can be improved. Yet the manufacturers aren't addressing themselves to this issue, I think, they ought to.

Mr. Bob Hynek: You say there's been a lot of pilot work done in that area, and there has been, many years ago. I think our feeling is that you are sure you can design equipment that will structural-ly handle the higher loads associated with higher RPM's, but it just doesn't make any sense econmical-ly to go any higher than 1,5, 1.6 with the present

diameter that we're talking about, because of the tremendous increase in energy that's required. You can do it better and more economically by other means, either more surface area. I think air in the system is being a major factor. It just doesn't make any sense to go that much higher because the energy requirements are going up at least as a function of two, two and a half power with RPM. Plus, probably a greater cost of equipment to handle the increase torques associated with it. So that isn't a direction that we, at least we, have elected to pursue as far as further improving the flexibility of the process.

Mr. Gary Davis: There are some benefits to higher speeds. We have operated at three RPM with a 11½ foot unit. It is more energy, but the levels of DO are much, much higher. The removal have decidedly improved. It is something that we would be very much interested in.

Mr. Bob Joost: The 1.6 which has been the criteria of most manufacturers, in order to get the organic removal, really doesn't hold true in all disc design. They were getting the organic transfer at .81 RPM on 11 foot diameter discs. We used 1.2 RPM. It's off standard. We got the organic removal. The 1.6 RPM came from people trying to get that same organic removal as given on a flat disc.

Mr. Bob Warren: I don't think increasing the rotational speed is the way to go on the RBC systems. They are, right now, the power requirements for RBC systems are graded and that was originally anticipated. And I think, competitively, if you begin to increase the rotational speed, the power is going to go up exponentially, and we should be looking for other ways to improve performance rather than increasing power requirement systems. Right now, they are beginning to approach power requirements of activated sludge systems, much closer than they are to trickling filter systems and we should be looking for ways to keep that speed down if possible. Things like adding supplemental air with

an RBC system to be able to reduce rotational speed as opposed to looking for ways to increase rotational speed.

Mr. Frank Viteck: I visited some RBC installations that are obviously overloaded, not just in the first stage but unit after unit and I am puzzled that they don't just arbitrarily increase the speed by changing the shafts and the belts to get the units running up to the RPM that it will allow. Because it would seem to be the easiest, first choice to get the plant out of trouble. Because, after all, when an operator is running a plant, maybe his problem is caused by the return of digester supernatant. Maybe it's because the plant wasn't designed properly. But, for whatever the cause, he has the option of getting more oxygen transfer by increasing the speed. And they don't seem to be picking up on that obvious avenue of escape from the immediate problem. I don't quite understand it. When I asked why they aren't doing anything about that, they kind of shrug their shoulders and say, "Well they didn't know they could do that".

Mr. Ed Opatken: Does anybody have any comment as to--You know I hear this and I hear somebody saying "Let's increase speed to increase oxygen transfer." Does anybody have any ideas how you're going to be able to handle the increased oxygen transfer? I mean it's fine to say it, you know, but if the objective is to increase oxygen transfer. Can we do it? Do we know the oxygen transfer at 1, 1.6 RPM? If we increase it, what's the hell of increasing it if we can't prove it, quantitatively. We've got to do it quantitatively to be effective, I think. Anybody have any ideas there?

Mr. Frank Viteck: I don't see that you have to reduce numbers. You ought to know that you're not going to have any oxygen if the speed is zero. And we all know that it's going to increase as the RPM goes up. So the operator's problem, then, is to get more oxygen transfer to meet an update rate that he's not able to satisfy. So if he can get a little more RPM--even 1.7, 1.8, 2.0 out of his $7\frac{1}{2}$

horse power motor, when he might only be drawing 3 or 3½, the mayor and the council and the commissioner of the sewer board are not going to give him hell for drawing a little more electricity out of the system. They're going to give him hell if he doesn't make the BOD requirement. So, if it was my ass that was on the line, running the plant, I'd burn all the electricity I had to and I'd get those things going a little faster as an immediate --maybe it's just a patch on a rat hole type of thing--but I'd just to get the oxygen transfer up--not knowing what the numbers are.

Marry Bergs: First of all, I want to disagree that you want to do it without numbers cause I think you should have numbers. The other thing is I've interviewed operators and they would like some flexibility as far as running things in both series and parallels. It would help them in quite a few instances. And the other thing is, you seem to have quite a bit of overloading like on first stages and hardly any load at all on the last stages. What about using two or three reactors in the first stage, and then only one or two in the second stage and and staging it down like that, rather than using more energy to speed it up.

Mr. Frank Viteck: I have to appologize for not remembering the name of the author that spoke yesterday about Fort Knox, but you'll recall that that discussion concerned 36 RBC's that were lined up 6 by 6. After the paper was concluded I asked the hypothetical question of "Well what would you do if you could take all 36 and pick them up and put them in a different geometric arrangement? And he said, "Well, I don't really know off hand. I'd have to take a look at it". I have the feeling myself, well you wouldn't just go and put it three by nine--in other words, nine across and three down. Because that man showed that the nitrification does not start until the third stage. But I think his paper clearly showed that he didn't need six stages, necessarily. But he's got 36 good pieces of equipment to work with there and if he had someway to rearranging the organic loading, such that he was starting

out with a more reasonable loading up front and tapering it off to keep the organic loading about constant, working on down the line, he'd probably put them down in a different arrangement than the six by six arrangement. For that reason, I don't necessarily see that you have to have supplemental air if you could pick them up and put them down in a different arrangement than six by six. I think that's what we're talking about here.

Twenty years ago, I was selling rotary distributors. Every job that I saw on a trickling filter, never had the second trickling filter the same size as the first. I guess the only reason for that was that the organic loading was reduced in the first trickling filter. You didn't have as much loading in the second so therefore you reduced the size--the diameter in order to keep the rocks about equally loaded. I think we've got to be doing something like that in the RBC's somehow or other.

Mr. Bob Warren: I want to separate a little the academic from practical. Then you have to separate the operation control over the design control. Certainly want to talk about having larger first stages. Once you build the system you don't want to change the shape of the first stage unless you have provisions to do that. But really the last thing I think you're going to tell an operator to do or should be telling an operator to do is to increase the rotational speed to increase performance. Number One--given the history of equipment failure, of shaft failure. If you tell an operator to increase the rotational speed, you're talking on the risk now of potentially having a shaft failure, because your system is now operating at a rotational speed that is certainly additional wear and tear on your system that you wouldn't have had originally. Yes you are going to increase your oxygen transfer efficiency, but you can do that in other ways that are not going to stress your equipment. And I think that making those type of practical decisions in the field would warrant that you would

not do anything that is going to decrease the potential equipment life that already has such a poor performance.

Mr. Ed Opatken: I don't want to touch that. We talk staging. Does anybody know why we stage? Certainly every system I've seen has been stages. Why do we stage it? No that's how they're installed. You don't have to stage them the way the manufacturer makes them the way I look at it. Does anybody know why we would stage.

Mr. Bob Hynek: There has been an evolution, I guess, at least as we see it. We started out a few years ago with a German technology that suggested some reasonably finite number of stages as being beneficial, both from a biological standpoint, and I suspect from the, in effect, mass transfer, standpoint. We certainly found over the years with testing and operating of plants that yeah, it doesn't make a great deal of sense to have a substantial number of stages. What it boils down to, at least as I understand it, is as is usually the base, a trade-off. If you're looking at relatively low degrees of treatment relatively high flow conditions, high substrate concentrations in the system where the reactions of kinetics is not very highly concentration dependent, it doesn't make much sense to stage, is basically what it boils down to. When you get down to attempting to achieve very low effluent concentrations of both ammonia and BOD, it does make some sense to accomplish some staging as a function of, in effect, achieving better residence time distribution and probably using some what less surface area at the low concentrations than you would have to if you did it all in a single stage reactor. I do think there's no question but what the trend, first of all as far as our design recommendations are concerned, we've been, in effect, Christmas Treeing the staging criteria for a good many years. I don't know that everybody has caught up with that, but that's certainly been our recommendation for about five or six years. For the most part. Then again, you got to

make that trade-off. Because as soon as you do that, you're probably going to get down to the point where you're going to use more surface area, as opposed to perhaps adding at least two or three stages, again if you're trying to get down to quite low effluent concentrations. I know some fairly specific single stage organic loading criteria that have evolved as a result of a lot of testing, a lot of experience, which I think gives us a pretty good idea about how far we ought to load organically a given stage and that works pretty well as a means of limiting the size of a stage if you're going to have any staging at all. But very clearly there's a trend or should be in the direction of probability applications anyway that will end up with only a single stage. And you can certainly accomplish carbon removal and nitrification in a single disc stage. There's no question about that. The question may be how much you're going to give up in overall efficiency for that perhaps, additional flexibility that you might have a whole system of very low substrate concentration, very thin biomass and what is in effect a single stage.

Mr. John Gastman: I'll stick my neck out, and feel somewhat shy and relatively young at this thing--but as see as Professor Gaudy point out earlier that there's a tremendous relationship between fixed film processes and mixed processes that I think we all can forget when we look at fixed film processes, and that, in part, because it's hard to measure, and that is essentially sludge age. It's a central parameter in activated sludge systems. As far as the microbiology goes, it's the same principle in a fixed film process, either anaerobic or aerobic and that you've got two conflicting criteria. One is how hard can you push the system in getting high removal rates and, inherently, that means you're going to have high growth rates, and the sludge age is essentially controlled by sluffing and you're going to start loading very highly, you're going to get a lot of sloughing, you're going to have low sludge age, even if you can't actually-even if it's very

difficult to measure. And you get low sludge ages, you get poor quality effluents or lower quality effluents--not necessarily poor, and you have the chance of reducing nitrification or eliminating nitrification. And so we've got to constantly keep in mind the objective and if the objective is a very low soluble effluent and a high degree of nitrification, you absolutely must have a significant sludge age on the, either in the activated sludge system or the fixed film system. And therefore, you can't just keep with a uniform loading throughout all the discs. And I think as the last speaker pointed out, there will be times in which you won't need that, in which you can push the loading rates much higher. I see this as one of the reasons trickling filters had a hard time in the early phases. The historical trend was you had a trickling filter built, the city grew, you had to push more water through the trickling filter, increasing loading rate and you lost the sludge age at the bottom of the filter and you reduced the effluent quality. Now it all comes plastic media and you can increase the sludge age again. But then the trend is to say, O.K., now that we've done that, we're getting these low bio-growths, why don't we find some way of pushing the system harder. And that's going to reduce the sludge age and we're going to be back in the same circle, if you don't keep these two conflicting ideas in mind.

Dr. A.F. Gaudy, Jr.: Thank you very much. I see we have a very spirited and active discussion on RBC's and we can come back to that later but Wes Eckenfelder is chomping at the bit to tell you something about trickling filters and at this time, well in a minute or so, I will turn this spot over to him.

Professor Wesley Eckenfelder: Thanks, Tony, We go over trickling filters which was brought up by my academic college. I'd like to make a few comments since I'm probably the oldest trickling filter guy in this room. And I'd like to say that of all biological processes engineered and opera-

ted by man, the trickling filter gotta be the oldest. That's not all biological processes, the oldest is the oxidation pond but they have been engineered by a bulldozer operator and have been operated by God, so. . . They do not fall in my category. You might say, when I first got out of college back in Tony's day here, my first job was at the treatment plant at Ridgewood, New Jersey, and they had the old fixed nozzle trickling filters, alternate dosing tanks. There's not a lot of this group who remembers those anymore, but one characteristic of those is that every night I went home--had to spend one hour picking the flies off my body. That is the one thing the plastic media did--got ride of the Psychoda flies. But also it got ride of the trickling filters. . . Those of you who have occasion to read the old, old literature have undoubtedly come across a number of papers I wrote on trickling filters. And when I was asked to discuss this topic here, I dug back into these old archives. And I found out the last paper I wrote on trickling filters was dated 1962. So I wondered about that, kind of looked into it, and what kind of happened is, and I'm glad I addressed another problem, industrial waste. But, with domestic sewage, in Public Law 92-500. Everybody got to get the BOD down to 10 down to 20, and the poor old trickling filter couldn't quite make it. So everybody built an activated sludge plant. And after they built the activated sludge plant they managed to get the BOD down to 40 and the suspended solids down to 60. And so as I speak to you here, we are not going back to the trickling filter. In fact, I used to put the trickling filter in my course under general topic, Historical footnotes, but now I've dusted off all those old archives and we're back in business. Now that brings us to the point of what are the important research needs that I see. And I want to address the kind of question you, as Ed did, but certainly don't want to restrict this to my own personal philosophies here. As to what do we have to look at now with the resurrection of the trickling filter. This topic would have been more appropriately discussed on Easter Sunday. However, we are

still fairly close to Easter, so I think it will be equally appropriate now.

Firstly, treating domestic waste water. I feel a major emphasis needs to be on developing, evaluating, looking at different filter packing configurations. And then we have to work at optimizing packings to optimizing performance. Personally, I think there's enough data and enough scale-up information and if you listened to my colleague this morning on our paper, you'll be partially convinced that it is possible from all the myriad information we have to be able to ascertain what are optimal packings. I really don't feel as far as domestic waste water goes, that we have to have a massive research program reinstated on the treatment of domestic waste water with trickling filters. Because I really think there's enough data, as long as we keep out the Psychoda flies, we should be in pretty good shape. And, as Ed did, I would invite a few comments on this whole packing deal here, because I think this is the key in a way as to handling domestic waste waters and trickling filters. Now I know some guy out there sells patents. He's gotta have a comment on this.

Mr. Kenneth Gray: We have a number of installations throughout the country and we basically have done these designs based on Schultz equation which is very similar to some of your work that you've done. And going back to those plants and correlating the operating data with that equation, it correlates very well. In fact, one of the plants area, Zansville, almost exactly predicts their actual effluent with that design equation. We're seeing some work being done, not only looking at the --not necessarily the packing but the clarifier designs--may be possible to go into a little deeper depths. Typically most of the clarifier designs about 10, 12 foot sidewalls.

There is some work being done by Brown and Cladwell that suggests that maybe something in the range of 60 feet or something may give you better

performance at relatively low operational costs.

Shelton Roe: We'll agree with it that Brown and Caldwell is being a real advancement in the trickling filter and then especially as far as getting low effluent BOD. We have some European experience too where you and you know sometimes we feel that we get some extra efficiency, but, over there the depth of a trickling filter is much shallower than the design here. Twelve feet is a normal.

Craig Edwards: I feel that I have to make an arbitrary comment here because I'm the manufacturer's representative. I think one of the things that Microfloc feels is depth innovation of the trickling filter time of the recirculating solids to the top of the tower, thereby, what we call activating the tower. We feel that, in numerous studies, that's been demonstrated to increase the tower performance. And, it then follows from that, we feel that we could return solid concentrations to the top of the tower. It is extremely important to have a media which can withstand those kinds of solids load problems. And based on that will be the configuration which we represent. In terms of the design equation, basically we use something similar to the McKinny equation for the design and we also are able to demonstrate the correlation of plant performance with hydraulic load.

Dr. Wesley Eckenfelder: O.K. I will move on to the subject of industrial wastes. I'm probably going to introduce a couple of controversial topics here, which will get all the academics on me at least if not anybody else. And that is, firstly, from the analysis that I have done, and this was like about five years ago, indicated that a trickling filter was most effective for relatively high strength industrial wastes for about a 50 percent BOD removal. And that if you attempted to carry a trickling filter to 90, 95 percent, it was not economically cost effective. At least that's the information I looked at at that time. And one of the reasons, of course that mitigate toward trickling filters now, and, again, I go back to my old archives data here.

What we did at one time to evaluate to a trickling filter, we were achieving upwards of 16 pounds of oxygen per horsepower hour on a comparable activated sludge basis, so obvious power economy. The questions I raise here, and since the topic of this Workshop is "Where should we direct our researches", is that, if you examine biological reaction kinetics, it would appear to me that as I go through a trickling filter, if I have a mixture of organics, that I would tend to take out the easy stuff first. And if you examine this, the implication becomes that your overall organic removal rate--if you go from the filter to activated sludge--materially goes down. Now the further implication of this case A is that if I have to get down to a very low effluent, a 10 say, soluble out of the activated sludge, that I gain no substantial decrease in my aeration tank volume because of the reduction in the organically remove rate called coefficient, mathematically. But, of course, I would gain substantial reductions in power.

By cite case B, the indication would be there that by virtue of the fact that I have taken out a certain category of organics through a trickling filter, that I would then develop a different kind of population in the activated sludge process, following which would be more effective, taking out the more refractory organics than would otherwise be the case.

I cite these two cases having absolutely no data to back 'em up. In this instance, as I suspect that nobody else has any data either, I cannot be proven wrong, incorrect, or screw up. But I do feel, and I'd like to open this question to the group, that in the industrial waste case, the most immediately probable use for the trickling filter is in combination with the activated sludge process. And there's one other reason for this--that virtually all industry today in view of Public Law 92-500, has an activated sludge plan out there, sitting there. It's not going to abandon them. The question is, how do you tie

these together to get the most correct effective answer. I guess I've indicated a couple of possible scenarios here but, at least from what I can see, the data is not available now to be able to make those kind of engineering decisions. Maybe it is, and that's the reason I open this up for some comments or questions from the group.

Mr. Ed Opatken: I've heard this statement before. That the K value for the initial organics that are removed is because it's easier to remove. They aren't refractory. Has anybody ever taken that particular position and ran a study to see how that removal rate is and then take the same waste, dilute it by a fourth, and see if you get the same removal rate, to find out whether it's concentration dependent or is it reaction rate dependent?

Mr. Orval Matteson: This is perhaps the time to make my comment because you, sir, have introduced the, I think, very valid point that a tremendous of waste treatment systems in the United States today are activated sludge. If we're not going to get good at the activated sludge, to go to something else while we have it. But my suggestion is that the solution to the problem is to try to combine within the existing facilities that you have--in this case I'm going to talk about activated sludge. That's why I came to this Conference, primarily was to suggest a procedure of combining within the activated sludge system those beneficial aspects of all these other types of submerged biological filters. Not the same physical structure, but the same bio-chemical results. I contend this can be done for very little money. And I think this is the challenge you've got here--really to upgrade waste water treatment in the United States in the existing systems, as well as to improve systems that have yet to be constructed. I'm very glad to bring about this fact other than suggesting some way that you could put an RBC in an aeration tank and somebody has to prove to me that that's the cheapest way it can be done with discs. But the reason for doing it, primarily, I think, is not to increase the aeration, but to increase the en-

vironmental condition that Dr. Gaudy spoke of. Waste water systems, as I said before this organization and this audience, are really not pumps and tanks and media, and RBC's. They're bullets. And to me, this is the challenge, how do you update the existing system in the United States without putting a great deal of money into it and a lot of construction?

Mr. John Wolfram: I haven't done work on what Mr. Opatken has. I've done a comparison between a one-fourth dilution with recycle versus one-fourth dilution with tap water and the results were dramatically different. The recycle proved at least 30 to 50 percent better by the nature of recycling the enzymes and so forth. Now as to the suggestion by Professor Eckenfelder here on the 50 percent removal, I think the proof is in the load approach, the curve where you plot one over F over M versus on the vertical scale the log of removal rate. You do get, quite frequently, a break in the curve. And this break indicates that while the curve is steep, you have a very good substrate removal rate; and a very good utilization of the volume while after it breaks and becomes sort of parallel to the X axis, you get a very poor utilization of your volume and you should stop at that point and operate above. That is in the much higher loadings. And these higher loadings that provide the efficient new utilization of the volume will yield your 50 percent removal within network. But I would like to address modeling question which was brought up by B.F. Goodrich, that the remodeling built by depth or time and so forth. I think we're again forgetting that we're dealing with a biological system and we should model load per square centimeter, or meter or surface area wetted and well slimed surface so we have a direct relationship to what the bugs really see. And not the arbitrary time versus depth. And one other point I'd like to make is the recycle issue, in case of industrial waste which is usually the subject of a lot of controversial statements. With certain wastes, the recycle is to even 400 percent are extremely bene-

ficial, while in others they are not and we've worked on a case where we've had six biofilters in series--six stages and five pumping stages. It was really a triumph of engineering over common sense, in this respect. And that particular operation with 120 feet of continuous media yet yielded 35 percent BOD removal in one through shot while a tower, of just one sixth of that depth, with a 400 percent recycle, did provide something in the order of 60 to 80 percent BOD removal, so there is a trade-off of pumping expense and power for pumping, a relatively low depth filter versus high depth and one through pumping.

Mr. Roger Ward: Your suggestion of up-grading activated sludge if a fixed filter process has been done very successfully. And I might add that it makes so much good sense, I can't understand why it isn't being done more often. The actual power savings to get that easy BOD out with a trickling filter or a fixed film process is about ten to one in favor of the fixed film process. The capital costs are a really big thing. It seems to me that fixed film processes are so efficient at getting the easy BOD out that they should be the logical choice.

Mr. Wesley Eckenfelder: I would agree with that. We have one more. Then I'm going to move on to my last topic because Art Freedman here is getting hungry so we got to get him on the program here too.

Mr. James O'Shaughnessy: There have been a number of issues that have been raised. And so far academia has come forth like a gallant knight to call. I'd like to address some of these things. To come back to Ed's first question, why we haven't come in the forefront to the RBC problem. If you look at it, I think our track record indicates that we'll let the regulatory agencies, the vendors, and the plant operators come up with the failures and then we'll put out what we can do. And I think we can point out very proudly to that record.

Today, if you'll look at the textbooks that we're using and all the programs, no one has a good

design process for RBC's --so let's go on to some of the other problems such as kinetics. When we get to dilutions, do we find what's needed? We had a study where we had a total dilution in industrial waste that gave us 95 percent removal as compared to 60 percent on the straight waste. Again I don't think we can say industrial waste unless we can say everything's waste dependent. There are toxins and other things. But in terms of fixed film reactors, there's a point I'd like to bring up and since we're talking about fixed film reactors especially talking about trickling aerobic fixed film reactors, after three days at the Conference, we come down and everyone is working on their model or we're presenting how good our removal was compared to someone else's removal. What we're forgetting is what we have to do is address what we're not removing. If we're going to make any progress, it's what's left. And in a fixed film reactor, everyone can do wonders with the soluble portion, and we just leave the residual particulates, especially in the laboratory in the university where we have a nice little reactor, which you want to make sure you have a soluble waste, even if it's synthetic, so you can get a nice little paper and say, "This model is going to work." That isn't going to translate to the real world, and that's the way I think we have to get to make any progress with frictional reactors.

Mr. Hallvard Odegard: I'd like to comment on this last one because I really have felt on this Conference too that you have been talking or we have been talking about the biological part of the process, but you must always remember that to make the good water, you have to separate the particles. And there have been no, at least very few, comments on that. And, at least, the problem with us is such that if you have a very high loading on the trickling filter or the RBC, you get a bad settling floc, whether you get a one part of the floc are settled very well and then you get a pinpoint floc that can't settle. And that's why we, as I told in my paper today, we combine this with phosphorus removal and then we have this combined precipitation. And

if you're not interested in that, you can, of course, increase the floc settling problems in other ways.

But I am also--I think at least that we will be looking very closely into other ways of separating the floc. We have been working with filter clothes, strain assistance, and things like that. And they are not so dependent on what kind of floc you get. In the Norwegian design criteria called trickling filter with combined precipitation, we allow three times higher organic load if we use chemical participation in addition, or in combination, with the biological treatment. So if you can overcome this oxygen problem, which you can by means of, for instance, aerating the thing, you can go up to much higher loads than, at least, is done in the U. S.

Mr. Wesley Eckenfelder: That brings me to my last point and it kind of ties in here, which I refer to as maximum concentration limits on a trickling filter. I'll cite one example here which will speak for itself. The city of Allentown, Pennsylvania, which put in a trickling filter plastic pack treatment unit for a couple of industries, a brewery, a food industry, 1976. The filter operated, I'm told for a period of five months before they had to evacuate the entire county. There was an odor problem and it's presently under all kinds of litigation, but I think the impact of concentration organic loading on such things as solids-liquids separation, odors--all these factors, to me has not been adequately addressed. And to tell such guidelines as for where you end and where you begin--I think we need those guidelines if we are going to aggressively use trickling filters for high concentrations of industrial wastes. On that with this I'd like to turn it over to my co-speaker, Art Freedman, who will tackle the underwater bugs as opposed to the above water bugs.

Mr. Art Freidman: I want to thank my great friends who have known me for many years as Art Friedman. Jerry Grady over here, Max whats-his-name at the

other end. I guess when everything else fails, at least from organic overloading from what I've been hearing, we end up in my segment of the program. I don't know why I'm a defender for submerged anaerobic systems. I guess I've stunk up a few laboratories and a few other places. But as I look around this room, people that I know that are working in the area, I can say there's a large group of stinkers right over there too, and some more spotted around.

It appears to me that when we approach these limits that Professor Eckenfelder was talking about, we approach the age where we're starting to say, as we did with RBC's ten years ago, let's take a look at these new cure it all set of processes, these anaerobic processes. Jim Young started us off back in the mid sixties. There's been some good work since then and it's beginning to pick up. Let me take a little different perspective here. I don't have any good stories. I want to challenge you with some areas that I think need some more research and hear your feedbacks. Right now, most of the anaerobic processes are pretty much black box approaches. We really don't know very much about the fundamental microbiology that's involved in these processes. Such things as synergism and competition, mutualism, what's toxic to these particular microbes. How we can go about forcing them to attach to surfaces. And the other side of that, what causes them to slough off? We have some ideas and we can make some gross guestimates but it really turns out to be a black box approach.

Another area that we know very little about in terms of these areas is how to design a system with an adequate safety factor. If I go out and start talking to potential people about the design of a system, the question really is, "How reliable is it?". There are many, many people who would spend 25 or 30 percent more on the initial capital cost if they were sure that that system would still be functioning well ten years down the road. I'm not sure that we can make that statement. At least with the little trickling filter, I know what I

bought. With an activated sludge system, I prayed for what I bought. With an RBC system, sometimes we're up, sometimes we're down. The one thing I can assure you is with any good functioning anaerobic system, we're just going to be a bunch of stinkers, no matter what. But these can work and we're having some real problems. Before starting with the fundamental microbiology, such things as particle separation on the outside. You know, I'm great academically, draw lots of lines on flow schemes and there's always a set of arrows that go off the bottom of the blackboard. And I tell the class, "We'll cover that later in the semester. That's sludge handling." No one really addressed these problems satisfactorily with the anaerobic systems. And we need to know more about that; in my opinion. We have real temperature problems. Oh, it's real nice and we all know we should run at 35 degrees centigrade. But can we run at other temperatures equally successfully? Birch does some good work in that area.

The answer appears to be yes but can I start a system up under those conditions? I know that sooner or later, no matter what the process, if it's biological, I'm going to have a failure of some kind. Then I'm sitting there and I say, "How do I restart"? And unfortunately, with my kind of luck, everybody here who knows me, my good friends included, it's going to happen when the time and the temperature is going to be working against me for one reason or another. It just always seems to work out that way.

We've not addressed questions concerning nutrients in the anaerobic systems. We'd like to ignore those questions. We don't know much about fundamental things in terms of say the mass transport questions that we need to address for some of the media that have been proposed for anaerobic systems. Not only that, we don't know much about the transport of intermediates between one group of these anaerobic microbes and another. There are lots of challenging problems out there, and I'd like to have some of my friends whom I'll call stinkers

pitch in and help share some of the potential solutions or some of the problems they see.

At this point, I want to turn it over to you.

Mr. John Vitek: Again, although I don't recall the author's name, I did hear the paper this afternoon discussing anaerobic digestion at 35 to 37 degrees centigrade, and the comment made that, well, the methane was generated. I couldn't understand why no discussion was made about what would appear to be the ease of taking that up to the thermophilic range of 125 to 130 degree Fahrenheit and if all that methane is available. It seems as though all the R & D work and investigation is going on with the so called mesophilic bugs, why aren't you guys in the university looking at the thermophilic? I don't understand that.

Mr. L. Van den Berg: We have had a little bit of experience with thermophilic work. And, to make a long story short, we cannot raise loaded rates above what we can as mesophilic. Start up problems are bigger. And so are stability problems. And this, of course, all related to fixed-film reactors. I think a lot of the work done in CSTR's is somewhat different because the thermophilic bugs grow faster than the mesophilic. And it may be possible to get somewhat higher loading rates and more stability in CSTR's mesophilic but certainly the fixed-film work that we have done, there is nothing in there that would make us choose thermophilic over mesophilic except, maybe, if you have a waste that's already at 140 so that you have a problem cooling it down.

Mr. John Vitek: I'm not familiar with the CSTR nomenclature. What's that stand for?

Mr. L. Van den Berg: That's the complete ~~disturbed~~ tank reactor, in other words, a normally mixed reactor without any retention of biomass

Mr. John Eastman: I was going to comment a little bit on sludge, a different question but it turns

out to be very related to your last question, that you're posing. And that is, I think in many of these anaerobic physical processes we've got to be very careful about looking at what happens in both the vault liquid, as well as what's happening in the film. And, in fact, there was one paper today in which some data was presented which indicated that--it was a highly soluble, readily degradable waste--largely sugar content. That indicated that with a fairly long liquid residence time, they ended up with some very high suspended solids in the effluent which they did not get at low hydraulic residence times. And, to a large extent, we've been talking afterwards, some of us thought that we were getting a tremendous growth of acid forming organisms in the liquid and the methanogens were on the fixed film. Whereas in the short hydraulic residence time, we were getting all the processes taking place largely on the film. And I've seen some evidence to maybe indicated that maybe under times in which the acid forms is controlled very rapidly and have essentially a very short sludge residence time which get very high yield coefficient with quite a bit of solids coming out of the system. And a lot more work needs to be done on looking at the acid forming organisms in the hydrolysis processes, where they're occurring, and the relationship between the yield organisms and the residence times, the environment which organisms grow that time.

Mr. Ed McCarthy: With regard to yields now. We've done some data which is, I'm afraid to say, proprietary, but that's the way things are in industry most of the time. And we don't want to tell you, mostly because we're afraid of exposing our sins, other than anything else. But basically, we're finding reductions from aerobic biological sludge at fairly high sludge ages--and I'm saying like 20 to 50 days--a comparison between that and yield in fixed filters. Fixed anaerobic filters. A reduction effect of 10 to 20, a factor--not 10 to 20 percent--a factor of 10 to 20. These are reproducible, gotten over a period of years not months. The second thing is the major impetus from my viewpoint, going to-

wards anaerobic systems is really solids handling. Those of us who have to work for industries know what a mess we're in right now with solids handling. There are more, I don't know what you want to call them, anaerobic holes in the ground around the country than most of us care to admit. They're there. The final point with regard to something somebody mentioned earlier, I think it was Ed Opatken, did I get his name right? mentioned somebody in was it Fort Knot or somewhere had 36 RBC's. I don't know how he manges, but we've got one plant, which due to a series of expansions, has nine parallel trains of activated sludge systems. And that is the biggest single disaster that I've ever seen in my life. I've tried to operate it for a while. Y u cannot possibly distribute flow to nine parallel systems equally. I mean some of the clarifiers in those systems are anaerobic year round. It's Why would anybody want to do something like that? I've heard the plan of engineering over common sense, but that's ridiculous. Does anybody know how to distrubute flow to 36 systems?

Mr. Henry Nelson: I'd like to pick you up on one point, Al. I can't do anything but agree with your suggestions, but it surprises me that you didn't come out with probably the most obvious research needed at this point in time, and that is something that--Oh perhaps I should introduce myself. Maybe we have a slightly different point of view from across the border, but permit me to make a few observations from afar. One of the problems that we're faced with is that we live in the here and now and people come along, consultants come along, industrialist come along and say, "How do we install an anaerobic filter? How do we install an anaerobic process? We'd like to recover energy from our wastes. What do we do? How do we go about it?" I put forward to you that maybe we should be doing, as well as attacking some of these more fundamental issues, we should also be trying to satisfy the needs of here and now, provide process design criteria. I'd like to think that we're making a little bit of a contribution in this area by examining four parallel systems of four different types of process

configurations at our laboratories. Nevertheless, this is pretty slow work. Anaerobic systems do not respond as quickly as aerobic systems. This kind of data generation is slow and it's difficult. And I think more attention should be paid in this particular area.

I sat here quite patiently listening to the three majors areas--the RBC's, the trickling filters, and the anaerobic systems. And it strikes me that you could pass a few general comments right across the board, in terms of the areas which you should be addressing, in terms of film systems in general. Our friend from the Michigan State University--well I'm from across the boarder. Michigan is a little bit foreign to me. Ah, I think he had a very valid point which absolutely, I'm surprised that you didn't pick it up either this concept to this SRT control, you made the point that trickling filter applications died with the introduction of PL 92-500. PL 92-500 and its successor, The Clean Water Act, also define a series of trace organics that you have to remove. No body at the meeting has mentioned trace organics, essentially, the whole week. And it strikes me that why is in nobody is generating information to satisfy the legislation. The most obvious way to go about it from our experience in house and out in the field is to execute SRT control. I speak only from activated sludge experience but also biological fluidized bed systems. I don't see terribly many people defining their systems in terms of SRT control. Also, I don't know how your energy costs are escalating on this side of the border.I come from overseas before I came into Canada. Energy costs were upermost in everybody's minds and some of the most obvious approaches you should be taking is trying to define which of these sytems are, in fact, efficient in terms of energy cost savings. Very few people seem to be defining the relative success of their operation in those terms. Another issue which for us in pretty critical in terms of input to the Great Lakes is improved along. Time and time and time again, people come forward at conferences--last year's WPCF Conference --a lot of

effort was made to project O and M papers. We don't really see enough flexibility built into basic designs for trickling filters, RBC's. Anaerobic systems you can't criticize at this point in time because they're still in their infancy and we're still struggling as we're going along. I think a lot more effort should be put into generating as flexible an operation as possible by improving process control systems. Somebody earlier on, I think it was our Norwegian friend, made the point that your system is only as successful--it doesn't matter what you've got at the front end--whether it be activated sludge, RBC, trickling filter, or whatever, you're only as good as your settling system. OK? Your clarifier is either going to kill you or make you successful. I think the two systems, both your biological and your clarification steps should be considered as an integral system. They shouldn't be divided as one being separate from the other. That's another issue I think should be addressed--right across the board. I guess the final item I've got--no hold on, two items--use of modeling, again somebody mentioned something about, I think it was John, modeling systems, making a comment during a break this morning. Dare I put my neck on the block with Wes sitting up there? I think models are fine. OK? You can model any God damn system you wish. But when you're going to invest anywhere between five and twenty five million dollars in a waste water treatment plant, there is no one in his right mind who's going to put a design together based on the model without doing any treatability work. I don't think model systems are sophisticated enough nor good enough, at this point in time, to base designs. I think treatability work just has to be a part of all this work. And the final comment I've got is in terms of utilizing foreign technology. Wes made the point, well how can we implement trickling filters in upgrading activated sludge plants? I think if you look and see what's been done overseas. The sort of examples I've got in mind are the ICI Lab out in Brixton in England was pushing its flow core packing in the early '70's. They were upgrading plants by sticking in high rate

plastic media filters ahead of existing activated sludge plants. I don't understand. We're not re-investing the wheel here. This is old-hat stuff. I don't see anything wrong with picking up what is being done overseas. We don't need to be ashamed that it is not true American or North American sort of know-how. Hell, if somebody has done it, paid the development costs -- let's use it, it is free. My final comment, and I hope Wes is going to come to the challenge here, is that he published a paper recently where he made a tremendous effort in integrating the biological systems under so-called unified biological theory. I hope that he is going to come forward with a few comments where he tries to tie these filter systems, some kind of integral approach, and identify these areas across the board which are commonly needed. I guess I've taken up enough of your time. Thank you.

Dr. C. P. Leslie Grady, Jr.

I was sitting here thinking about one thing which has been running through my mind all evening, and that is that every one of the research needs which has been spoken of has been very now, very practical in sense and I think that this is very appropriate for a meeting of this type. Also, I have been thinking that a lot of the research effort right now is focused on the now, unfortunately. I am not bright enough to know what the problems are going to be 15 or 20 years from now. I have a very basic belief that the solutions to those problems are going to come only through a good base in fundamental research right now. We have talked about liquid solids separation-- yet, we know very little about the chemical physical factors with the microorganisms that are involved in a block formation. We're talking about fixed films here. Where is the majority of work being done now on attachment mechanisms in fixed films? It is being done by dental scientists. Surely, you know that we who are having the biggest problems with fixed films should be in the forefront of looking at attachment mechanisms of micro-

organisms to surfaces. The problem was brought up with the particular organic compounds. In fixed film reactors, we have for the first time a very real problem present -- metabolic control mechanism within organisms. Tony and I looked at this problem 15 years ago and found that it was not important in an activated sludge system. However, this problem could be quite important in fixed film reactors. Because we do have regions within fixed films where growth rates are fast enough that the basic metabolic controls food, substrates are removed within these systems. We need to know a lot more about these things in order to predict what is going to happen to recalcitrant and other difficult to degrade compounds. It seems to me that there is a whole host of things which we need to be looking at from a very fundamental standpoint. Even though I can't tell you how we are going to use these things, I have a basic belief that if we do not look at these, we are going to be found wanting within the system when problems crop up. As far as integrating systems are concerned, I guess I have always believed that there is a unity to all of these systems that lies within the microbiology of the systems. Microbiologists are not interested in the types of problems which we have to deal with. As engineers, it is up to us to be doing the types of applied microbiology work that is going to lead us to better answers. When we try to model a system in order to understand it better, we need to know more of the physical characteristics. For instance, oxygen transfer coefficient and these sorts of things, and I do not see this type of work progressing very rapidly. I am very sympathetic to the here and now, and I agree that real world problems exist and that we need a lot of practical research. However, I feel that it is important that the record of this symposium show that there is a very real need out there for further fundamental research. And, hopefully, there will be some of us who will continue to do that.

Dr. A. F. Gaudy, Jr.

Thank you very much. In case anyone was wondering what Les Gray was doing up here and didn't have an announced role -now you know why he was here. Les has hit a few nails right on the head. And, I think the gentleman from across the border -I assume it was Canada, not Mexico- did a great job of summarizing. I did not take full notes on this because I was going to give Les a chance to make the comments he made. One comment he made was that we should consider the settling tank and the clarifier the same. I think that this needs a little more interpretation and discussion. I feel that we should consider these as two related but separate unit processes or unit operations. Most of the models and most of the things which have been done to understand the biological processes have to do with biochemistry which is related to the physical processes of agglomeration, detachment and separation. But, as the gentleman from Norway pointed out, we should not wed ourselves to clarification and sedimentation tanks. We should wed ourselves to the separation processes and all the things which go along with it - like chemical addition and a few other things like filtration. The minute we start to talk about, "well, we can take out the solubles, but the suspended goes out and therefore, we should throw the process out" - that's like killing the very, very useful golden goose. That is a separate process and we can take care of it by separate types of operations, realizing that these are related. This is my only opposition to what the gentleman said.

Bob Norcross

We are producers of antibiotics and microbes, and I have been laughed at by some of the best consultant firms in the country - that we have no real place. From what I have seen at this conference, everyone seems to focus on the microbiology. Our aim is that in the future, your engineers will begin to look at the very distinct possibility that there can be some biological

engineering in conjunction with the physical and chemical engineering that is going on at this point. And, I guess our real hope is that people will begin to look at what we are doing as seriously as some of the things done with pumps and tanks and reactors and so on and so forth. We have been at it for 20 years, and we are ready to make some progress. However, it is going to take your help and cooperation to find out whether or not it really works. In talking with Dr. Wu, it is our hope that when this night convenes again in two years, maybe some one will take a look at what we are doing and come up with some good answers which will enhance these processes and bring us along from a real art into a science.

Dr. A. F. Gaudy, Jr.

Thank you very much. This is incidentally a field that does need much exploration and research and there may be many things that one of the pollution control engineers with a biological bent and operators with a biological bent can do - to grow our own rather than buy. There will be times down the line in which we can perform ecological engineering to control those things ourselves. Basically, all biological treatment plants are biochemical plants - not only plants that treat the waste, but plants in which we manufacture the biochemicals that treat the wastes. I feel along with what the gentleman said and some of the remarks made by Les, and the convictions that many of us have, that our field will go forward much faster and on a much sounder ground when we realize that we are in the manufacturing process. We manufacture biochemicals which treat the wastes, and biological treatment turns out to be normally the most economical way to treat the waste because basically we make it as we go.

Dr. Yeun C. Wu

I would like to re-emphasize both Professor Grady's and Dr. Eastman's discussions made earlier! We have not yet fully understood the basic factors affecting the operation of fixed-film biological treatment processes. For instance, can cell attachment be accelerated by the surface modification of fixed-film media? If yes, what shall we do about it? The improvement of cell attachment rate certainly will reduce the system start-up time and, of course, it will make the system more attractive than suspended growth system. Thank you.

Dr. A. F. Gaudy, Jr.

It's a subject very basic to our field, and we will talk about attachments of microbes to surfaces, one microbe to another microbe surface other than biologically or chemically assisted fiber. That is one of the most basic questions in our field and yet, it is one of the least talked about. As we all know, many people have looked into it - Ross McKenny, Mark Tenney and a number of others. I myself have worked on it. I have never had quite the heart to recommend to a Ph.D. student that he make this a subject of his Ph.D. research because the poor guy would probably be working for the rest of his life to get his Ph.D. It is important to get to the problem. The thing that Yeun pointed out is what we should be looking at since we do not understand it. It would take a tremendous investment of brainpower, manpower and resources, and manual energy resources to translate to research funds which are not forthcoming in the current period. To understand that process is just like chlorination. Nobody knows how chlorination kills organisms, but we use an awful lot of chlorine. We just do not understand those things. That is why Universities should be doing these kinds of things. Universities need the backing from people in industry and from people in government. Universities should not be telling

people on the street necessarily, at least in my view, how to design; nor should they be providing the basic information for the equipment companies. They should be doing basic fundamental research and that is one of the basic fundamental things which needs to be researched. Before I turn it over to you, I'd like to say that the rest of our program tonight should be one in which we feel free to address ourselves to all three processes, to address questions and comments to ourselves and to our experts on the podium; and also to add to any of the processes anything in general to aid this very important process of biological treatment of wastes.

Dr. John Eastman

I hope that I am not stepping out of bounds here a third time. You have just brought forward something which is critical. Again, I hear a lot of discussion here and now for the immediate problems. I know that many industries have waste-treatment problems. I hope that some of these outfits would be willing to support some of the longer-term research, and not leave it entirely to the government to support the more fundamental basic research. Let us have a chance to work on the very real problem, and to have a little extra flexibility to include the longer-term goal. I feel that this would help everyone in the long run. As far as the short-term goals, it is going to be counter-productive. I realize that there will be much discussion here among the academics as to where we are going to turn for research support. It is not going to be forthcoming from the government. Therefore, if we do not find some means of doing research within the universities, we are going to sort of die away; and when we get back to look at these fundamental problems, it will take us a while to come back up. I hope that we can develop a cooperative arrangement -where I am not going to bide my time doing only theoretical problems. I personally do better

when I am looking at the fundamentals in direct relationship to some practical problems, rather than doing complete isolation - cooperation between academics and industry --both chemical companies and metal-plating plants-- which include the whole range of activities, a cooperative arrangement -not only in the sense of just dollars-of sitting down together, discussing problems and being a real partnership.

Frank Viteok

As a marketing manager for a manufacturer, I am afraid to say that you are going to be disappointed because, as you know, industry has been burned during the last few years trying to chase the ever-moving target in hopes that some profit would be there. Many companies are not here today that were very enthusiastic to make profits ten years ago. We have'nt seen the profits; nor do we really know what the future holds. I think it's a here and now, and look for the short-range at this point. Chasing the ever-elusive profit goal by long-range objective is not something that corporations are cozying up to. It is something that they are backing away from. It will be a good while before this situation changes in view of the high interest rates, the weak economy, and the changes that are taking place in the EPA. That is just the real world, and that is where we are today. All that has to bottom and come back again. I have been in the industry long enough to see times come and go, and right now it is kind of at the bottom. I know that you will be in industry a long time, probably after I have retired, but it's a difficult time now. We have seen some good-size companies come and go in the last couple of years.

Dr. A. F. Gaudy, Jr.

Thank you very much. Your statement is very true. The other gentleman's statement is also very true. The thing that seems to happen in our system is that most of the pressure seems to come from Washington, and Washington changes almost every four years. I have gone through periods where the Federal Government should participate and that was the word which came out of Washington and the Federal Government participated. The precedent set is that Washington is polluted and therefore, industry should spend the money to fix up, and therefore, industry should be grateful to fix up. Both of these attitudes are perfectly O.K. The trouble is that when we vacillate from one to the other every four years, we do not make much progress.

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