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UPGRADING ACTIVATED SLUDGE PROCESS WITH ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

The City of Crawfordsville, Indiana, is a community of approximately 13,000 residents and has a diverse industrial base. The major industrial wastewater sources are metal plating, printing, and wire fabrication operations. Wastewater flow from industry constitutes approximately 25% of the total plant flow.

The original wastewater treatment plant, constructed in 1940, was a 1.0 million gallon per day (MGD) primary and conventional activated sludge facility. Expansion projects through 1970 increased plant treatment capacity to 1.8 MGD by expansion of the primary and secondary tankage and aeration blower system capacity.

In 1977, with Federal and State financial assistance, the design of yet another improvement project commenced. Facility planning recommended the expansion of the average daily treatment capacity of the plant to 3.4 MGD and the incorporation of advanced wastewater treatment facilities (i.e., tertiary filtration). The most cost-effective approach for upgrading the existing activated sludge process to accommodate the increased

design organic loading was determined to be the installation of a fixed-film biological "roughing" (pretreatment) process so as to reduce the organic loading imposed on the existing activated sludge aeration tankage. On the basis of costs and operational flexibility, mechanically driven rotating biological contactors (RBC's), rather than trickling filters, were selected as the biological roughing process. Other additions and modifications to the facility included: expansion of the primary and secondary tankage, raw sewage pumping and aeration blower system capacity, addition of dual media filtration, dissolved air flotation thickening of waste activated sludge, and belt filter press dewatering of the anaerobically digested sludge. A schematic of the overall treatment process is presented on Figure 1 and related design data is summarized in Table I. RBC tank layout is shown on Figure 2.

This paper presents:

1. The design methodology used for the sizing and layout of the RBC units;

2. A comparison of the full-scale operational data collected since November, 1979 to the performance predicted by the design methodology;

3. The operational factors which affected the performance of the RBC units; and

4. The enhancements to the overall plant performance (e.g., nitrification and secondary clarification) which are attributed to the RBC process.

RBC DESIGN METHODOLOGY

The surface area requirement of the RBC's was generally based upon achieving a 50% reduction of the soluble five-day biochemical oxygen demand ($SBOD_5$) of the wastewater ahead of the activated sludge process. Design methods (published in RBC manufacturer catalogs prior to 1977) for determining the required surface area typically did not address the biological roughing application and did not consider applications for which the effluent $SBOD_5$ would intentionally exceed 25 mg/l. However, Antonie(1) proposed a design method and model equation for multiple stage RBC's that did address the cases for which the effluent $SBOD_5$ would be in excess of 25 mg/l. Figure 3 illustrates that design method, gives the model equation for the original sizing of the RBC's, and shows an example of predicting full-scale performance.

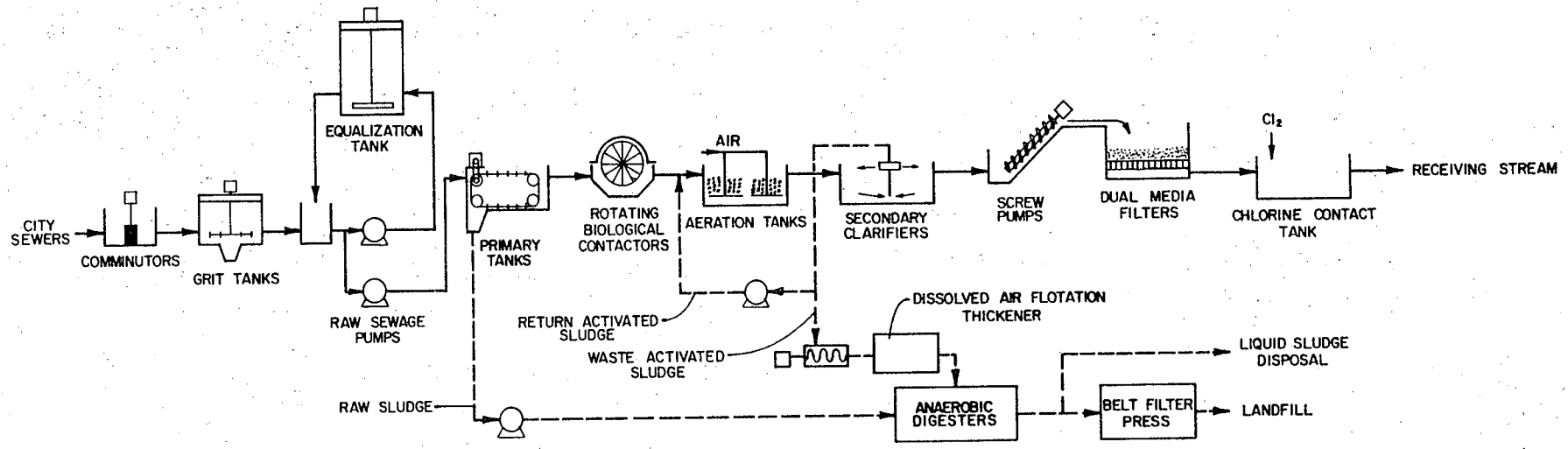

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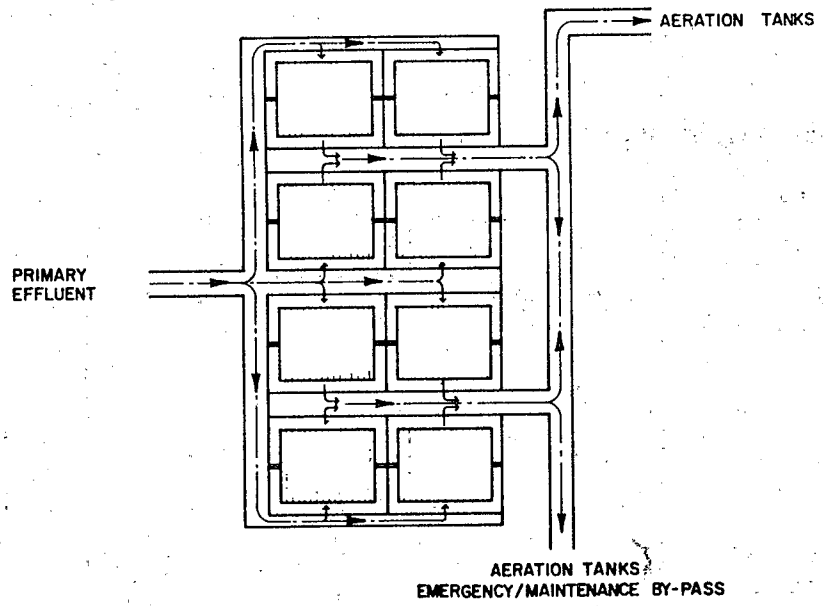
FIGURE I
PLANT PROCESS SCHEMATIC

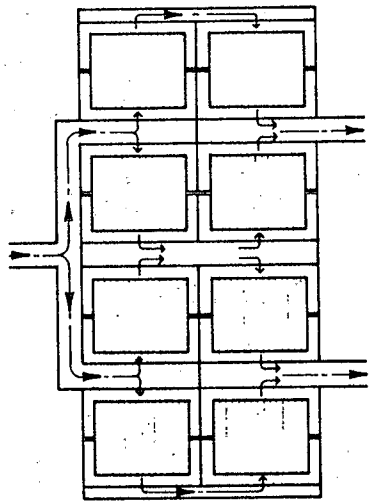
TABLE I

PLANT DESIGN DATA

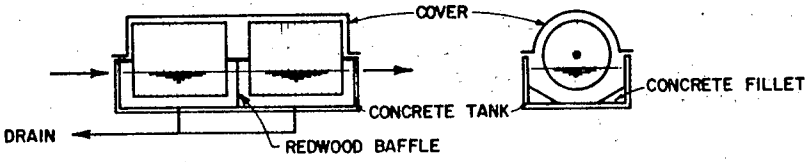
Raw Wastewater Strength	180 mg/1 BOD_5 (five-day biochemical oxygen demand) 190 mg/1 TSS (total suspended solids)
Raw Wastewater Flow	Average, 3.4 MGD (million gallons per day)
Flow Equalization Basin	650,000 gallon capacity
Raw Sewage Pumping	6 - 1400 GPM (gallons per minute) pumps
Primary Settling Tanks	8 - 10 Ft. x 35 Ft. units
Rotating Biological Contactors	4 - 96,000 Sq. Ft. units
Aeration Tanks	4 - 112,000 gallon tanks
Air Blowers	2250 CFM capacity
Secondary Settling Tanks	4 - 20 Ft. x 122 Ft. units
Filter Feed Pumps	2 - 2850 GPM pumps
Dual Media Gravity Filters	6 - 11.5 Ft. x 11.5 Ft. filters
Anaerobic Digesters	2 - 35,000 Cu. Ft. units
Dissolved Air Flotation Unit	1 - 35 Ft. diameter unit
Sludge Dewatering Belt Presses	2 - 2 meter units



PLAN VIEW
SINGLE STAGE FLOW PATTERN



PLAN VIEW
TWO STAGE FLOW PATTERN



SECTION OF TYPICAL
TANK LENGTH

SECTION OF TYPICAL
TANK WIDTH

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FIGURE 2
RBC TANK LAYOUT

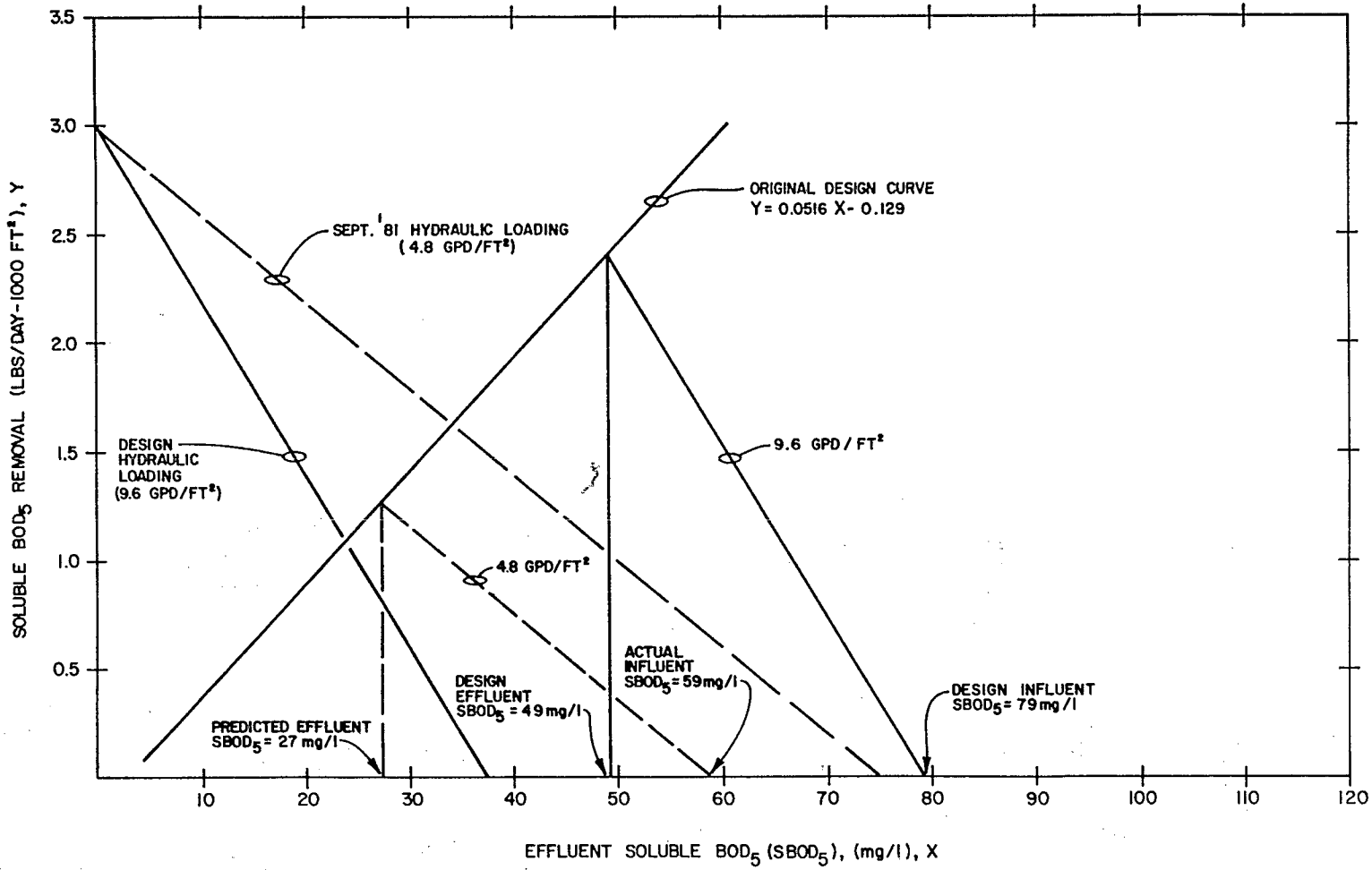

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FIGURE 3
DESIGN AND PREFORMANCE PREDICTION MODEL
FOR SINGLE STAGE RBC

The actual design technique involved several trial and error solutions. Four standard size 25 ft. x 12 ft. diameter (100,000 sq.ft. of media per unit) RBC units were determined to be necessary to meet the process requirements for 50% $SBOD_5$ reduction, if the RBC units were operated as a two-stage process. System flexibility was maximized by configuring the four units side-by-side and fabricating each unit with two distinct media sections. See Figure 2. This configuration permits the isolation of any one RBC unit (for maintenance) and either a single or two-stage process. Approximately 4% of the media is removed to affect the division of the RBC media into two distinct media sections. As a result, the actual surface area of the media of each RBC unit was 96,000 rather than 100,000 sq.ft. The actual total design surface area of media was 384,000 sq.ft.

The single stage design example is presented on Figure 3. The single stage, rather than the two-stage design example is presented because the RBC units were started up as a single stage process and have remained as a single stage process. The performance of the RBC units has not yet required a change to the two-stage process.

The final design effluent $SBOD_5$ for a single stage process was determined on Figure 3 as follows:

1. Calculate the design hydraulic loading.

   Design Flow = 3.4 MGD plus 10% for recycle streams = 3.7 MGD

   Design Media Surface Area = 384,000 sq.ft.

   Design Single Stage Hydraulic Loading = 3,700,000 gallons per day (gpd) ÷ 384,000 sq.ft. = 9.6 gpd/sq.ft.

2. Find the slope of the hydraulic loading line on Figure 3, at the design hydraulic loading of 9.6 gpd/sq.ft.

   The dimensions [gpd/sq.ft.] are equivalent to:

$$\frac{[\text{lbs. } SBOD_5 \text{ removed}/(\text{day} - 1000 \text{ sq.ft.})]}{[\text{mg/l } SBOD_5 \text{ removed}]}$$

   x 1,000/8.34 lbs./gal.

Therefore, to express the design hydraulic loading of 9.6 gpd/sq.ft. in terms of the slope dimensions on Figure 3, the hydraulic loading is multiplied by 8.34 lbs. per gal./1,000.

$$9.6 \text{ gpd/sq.ft.} \times \frac{8.34 \text{ lbs./gal.}}{1{,}000} =$$

$$\frac{0.08 \text{ lbs. } SBOD_5 \text{ Removed/(Day-1,000 sq.ft)}}{\text{mg/1 } SBOD_5 \text{ Removed}}$$

A simple graphical display of the calculated slope is determined by arbitrarily selecting the Y-coordinate as 3.0 and computing the X-coordinate:

$$\text{X-coordinate} = \frac{\text{Y-coordinate}}{\text{slope}}$$

$$= \frac{3.0}{0.08} = 37$$

Therefore, the slope of the design hydraulic loading line is arbitrarily shown on Figure 3 as intersecting the Y-axis at 3.0 and X-axis at 37.

3. Find the single stage design effluent $SBOD_5$, given the designed hydraulic loading rate of 9.6 gpd/sq.ft. and the design influent $SBOD_5$ as 79 mg/1.

   Shift the hydraulic loading line so that it intersects the X-axis at 79 mg/1. The X-coordinate of the point of intersection of the shifted hydraulic loading line with the design curve is the design effluent $SBOD_5$ which is shown as 49 mg/1.

Although a single stage configuration did not predict a 50% reduction in $SBOD_5$ at the design single stage hydraulic and organic loading, the two-stage configuration did predict a conformance with the basic design requirement.

EVALUATION OF DESIGN METHODOLOGY

Table II lists the monthly operating data of the RBC units, and contains two columns which readily indicate the

TABLE II

RBC OPERATING DATA

Month	Water Temp (°C)	Flow (MGD)	Hydraulic Loading (GPD/Ft^2)	Influent $SBOD_5$ (mg/l)	Effluent $SBOD_5$ (mg/l)	Predicted Effluent $SBOD_5$ (mg/l)	Actual Minus Predicted Effluent $SBOD_5$	$SBOD_5$ Loading (Lbs/Day-1000 Ft^2)	$SBOD_5$ Removal (Lbs/Day-1000 Ft^2)
Nov '79		1.75	4.6	77	39	34	- 5	2.9	1.4
Dec '79		1.70	4.4	91	56	39	-15	3.4	1.3
Jan '80		1.78	4.6	113	74	50	-24	4.4	1.5
Feb '80		1.72	4.5	94	57	41	-16	3.5	1.4
Mar '80		2.02	5.3	84	52	40	-12	3.7	1.4
Apr '80		2.23	5.8	74	43	37	- 6	3.6	1.5
May '80		2.40	6.3	116	67	59	- 8	6.0	2.6
Jun '80		1.71	4.5	95	46	41	- 5	3.5	1.8
Jul '80		1.78	4.6	109	54	48	- 6	4.2	2.1
Aug '80		1.91	5.0	83	43	38	- 5	3.4	1.7
Sep '80		1.89	4.9	74	40	34	- 6	3.0	1.4
Oct '80		1.86	4.8	98	52	44	- 8	4.0	1.9
Nov '80		1.82	4.7	100	55	44	-11	4.0	1.8
Dec '80		1.73	4.5	134	60	57	- 3	5.0	2.8
Jan '81	13.5	1.70	4.4	102	54	44	-10	3.8	1.8
Feb '81	12.0	2.43	6.3	119	63	61	- 2	6.3	3.0
Mar '81	13.0	2.25	5.9	109	58	44	- 4	5.3	2.5
Apr '81	16.0	2.47	6.4	82	46	43	- 3	4.4	1.9
May '81	16.0	3.28	8.5	75	44	45	+ 1	5.3	2.2
Jun '81	20.0	2.13	5.5	85	41	41	0	3.9	2.0
Jul '81	22.0	2.00	5.2	69	34	33	- 1	3.0	1.5
Aug '81	22.0	1.98	5.2	59	32	29	- 3	2.5	1.2
Sep '81	21.0	1.85	4.8	59	31	27	- 4	2.4	1.1
Oct '81	19.0	1.74	4.5	61	32	27	- 5	2.3	1.1
Nov '81	19.0	1.65	4.3	70	40	30	-10	2.5	1.1
Dec '81	17.0	1.81	4.7	69	43	31	-12	2.7	1.0
Design	17.5	3.70	9.6	79	49	-	-	6.3	2.4

NOTE: $SBOD_5$ data presented above were collected during single stage operation and are based upon 24-hour composite samples, each consisting of six flow proportional grab samples taken every four hours. Composite samples were normally taken Sunday through Thursday. Data listed above are monthly averages based upon a minimum of 18 daily composite samples per month.

validity of the design technique: predicted effluent $SBOD_5$ and actual minus predicted effluent $SBOD_5$. An example of how the predicted effluent $SBOD_5$ values were obtained is illustrated on Figure 3 for the September, 1981 data set. The slope of the hydraulic loading line was graphically depicted as follows:

1. Express the monthly average hydraulic loading of 4.8 gpd/sq.ft. in terms of the slope dimensions on Figure 3.

$$4.8 \text{ gpd/sq.ft.} \times \frac{8.34 \text{ lbs./gal.}}{1{,}000} = \frac{0.04 \text{ lbs. } SBOD_5 \text{ Removed/(Day - 1,000 sq.ft)}}{\text{mg/1 } SBOD_5 \text{ Removed}}$$

2. Arbitrarily select the Y-coordinate as 3.0 and compute the X-coordinate.

$$\text{X-coordinate} = \frac{3.0}{0.04} = 75$$

3. Draw a line with a slope equal to the hydraulic loading as intersecting the Y-axis at 3.0 and the X-axis at 75.

   The predicted effluent $SBOD_5$ value of 27 mg/1 was found by shifting the hydraulic loading line so that it intersects the X-axis at the monthly average influent $SBOD_5$ value of 59 mg/1. The X-coordinate of the point of intersection of the shifted hydraulic loading line with the design curve is the predicted effluent $SBOD_5$ value, which is shown as 27 mg/1.

In general, the actual performance of the RBC units has been reasonably close to performance predicted by the design methodology. The predicted effluent $SBOD_5$ values of the 1981 data set averaged approximately 90% of the actual effluent $SBOD_5$ values. December, 1979 to February, 1980, data are not considered representative of normal performance because the plant received industrial cyanide spills during December, 1979 and January, 1980.

RBC PERFORMANCE AND OPERATION

Figure 4 illustrates that the RBC units consistently removed 40 to 50% of the influent $SBOD_5$. The removal rates did not appear to be significantly affected by hydraulic or organic loading rates. For example: The average hydraulic loading rate in May, 1981, was 8.5 gpd/sq.ft. and the percent $SBOD_5$ removal for that month was 41%; whereas, the average hydraulic loading rate in November, 1981, was 4.3 gpd/sq.ft. and the percent $SBOD_5$ removal for that month was 43%.

Figure 5 is a plot of the effluent $SBOD_5$ versus $SBOD_5$ removal rate and the original design curve. Most of the 1981 data points are in close agreement with the original design curve. Better control of the sludge handling/treatment sludge recycle streams occurred in 1981 and is considered partly responsible for the improved performance of the RBC units. Also, the industrial source of cyanide was controlled.

In general, the performance during the summer months was better than during the winter months. Performance factors other than wastewater temperature probably account for difference in performance. The wastewater temperature does not fluctuate significantly (12.0 to 22.0°C). Furthermore, the coldest monthly average wastewater temperature (12°C) was recorded in February, 1981, but the performance during that month was a similar to the performance during August, 1981, when the highest monthly average wastewater temperature (22°C) was recorded. Plant operating personnel attribute the seasonal performance differences to an increase of sludge handling/treatment recycle streams (e.g., anaerobic digester supernate) during the winter months due to periodic interruptions of the disposal operations of liquid digested sludge and lower primary digester temperature.

The daily average $SBOD_5$ loading exceeded a RBC manufacturer's(2) recommended limit of 4.0 lbs. $SBOD_5$/(day-1000 sq.ft.) for 10 out of the 26 months of operation listed. Plant operating personnel daily have checked the bio-film of the RBC units for patches of white growth (which is a visual indication of undesireable forms of microbial life, presumed to be beggiatoa), and have seldom noted patches of white growth.

Routinely, the plant operating personnel exercise buried drain valves in order to assure valve operability. Significant amounts of sludge had been noted to be withdrawn during

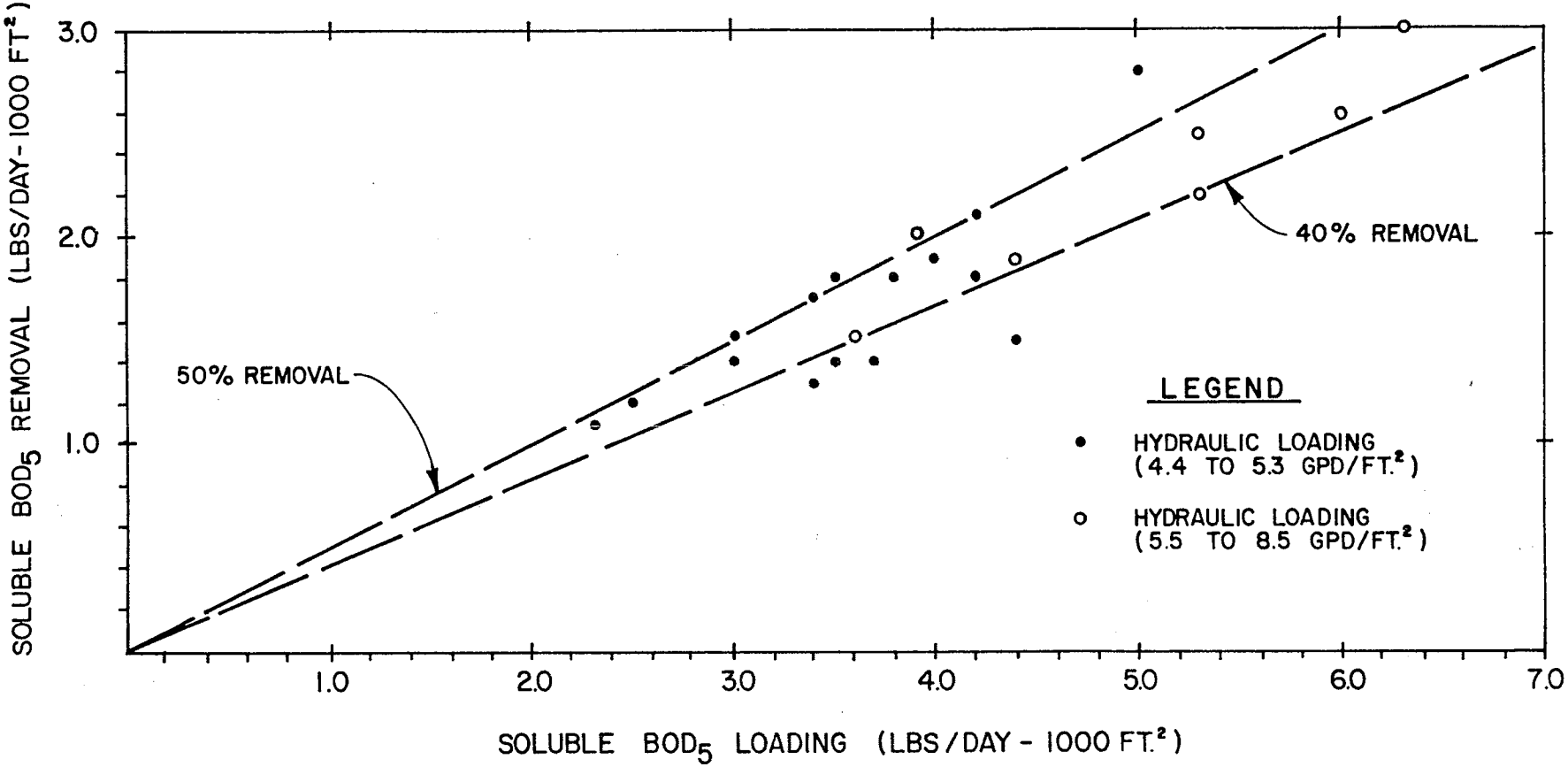

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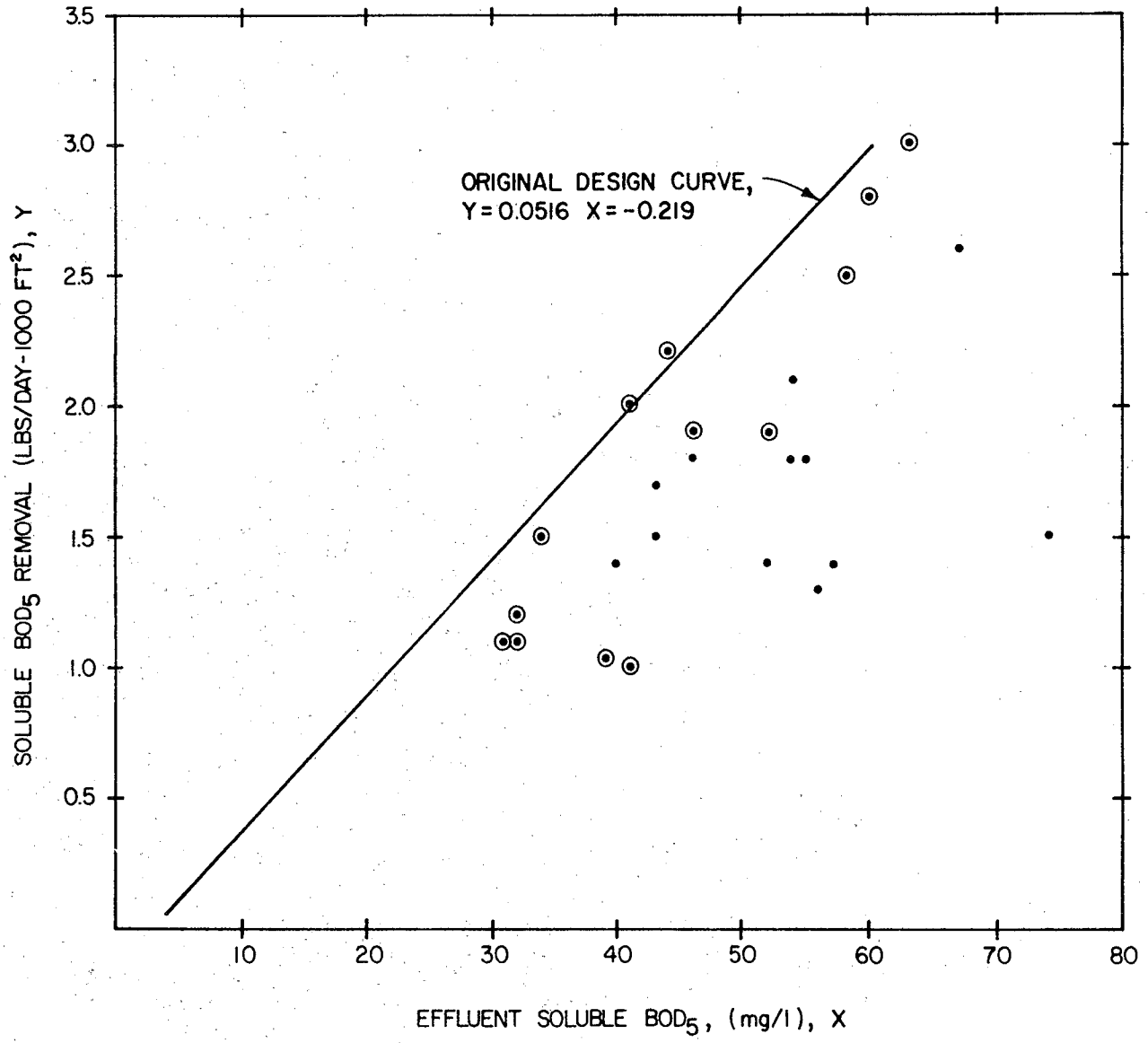
FIGURE 4
LOADING VS. REMOVAL RATES



LEGEND

⊙ OPERATING DATA FROM DEC. '80 THRU DEC. '81

• OPERATING DATA FROM DEC. '79 THRU NOV. '80

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FIGURE 5
COMPARISON OF ACTUAL OPERATING DATA
TO ORIGINAL DESIGN CURVE

the routine exercise of the RBC tank drain valves. Impressed by the amount of sludge withdrawn, the operating personnel decided to withdraw sludge from the RBC tanks on a regular basis. Coincidental with the decision to periodically withdraw sludge from the RBC tanks, the plant operating personnel also noticed that the white patchy growth on the bio-film appeared less frequently. Sludge is withdrawn routinely 2 to 3 times per week and more often if white patchy growth develops.

An attempt was made to correlate $SBOD_5$ removal efficiency with the practice of routine sludge withdrawal from the RBC tanks. Sludge was withdrawn from only two of the four RBC tanks during the period of November, 1981 through February, 1982. The daily average effluent $SBOD_5$ of the drained and undrained tanks was nearly identical during the trial period. Even though the data generated during the trial period did not support the assumption that periodic sludge withdrawal has a beneficial impact upon $SBOD_5$ removal, the plant operating personnel have maintained their periodic sludge withdrawal operation and have seldom observed white patchy growth.

AFFECT OF THE RBC PROCESS ON OVERALL PLANT PERFORMANCE

Table III lists the 1981 monthly operational data of the RBC and activated sludge processes. The secondary effluent typically was nitrified and had a total suspended solids concentration less than 20 mg/l. The biological roughing of the RBC process not only reduced the $SBOD_5$ load to the activated sludge process, but was also likely responsible for the excellent quality of the secondary effluent. Both the nitrification and the excellent secondary clarification that occurred would not have been anticipated for an activated sludge process which operated at an F/M ratio greater than 0.20/days, mean cell residence time less than 4.5 days, and a hydraulic detention time less than 4 hours.

In addition to $SBOD_5$ removal, the RBC process converted the form of the suspended solids entering the RBC's from a very non-descriptive particle type to noticeably long, dark, stringy biological solids. This formulation of biological solids ahead of aeration may be one of the enhancement factors that an RBC process lends to the activated sludge process.

TABLE III

RBC/ACTIVATED SLUDGE PROCESS
OPERATING DATA

Month	Water Temp (°C)	Flow (MGD)	RBC Hydraulic Loading (GPD/Ft^2)	RBC $SBOD_5$ Loading ($lbs/Day-1000\ Ft^2$)	Aeration $SBOD_5$ Loading ($Lbs/Day-1000\ Ft^3$)	F:M Ratio (Lbs/Day BOD_5 Per Lbs MLSS)	Aeration Hydraulic Detention Time (Hrs.)	Mean Cell Residence Time (Days)	Secondary Effluent TSS (mg/l)	Secondary Effluent NH_3-N (mg/l)
Jan '81	13.5	1.70	4.4	3.8	12.8	0.20	6.4	NA	35	1.4
Feb '81	12.0	2.43	6.3	6.3	21.3	0.25	4.6	NA	42	2.4
Mar '81	13.0	2.25	5.9	5.3	18.1	0.26	4.8	NA	20	<0.1
Apr '81	16.0	2.47	6.4	4.4	15.8	0.20	4.4	NA	11	0.2
May '81	16.0	3.28	8.5	5.3	20.1	0.24	3.3	NA	9	< 0.1
Jun '81	20.0	2.13	5.5	3.9	12.1	0.16	5.1	NA	11	< 0.1
Jul '81	22.0	2.00	5.2	3.0	9.5	0.11	5.4	7.5	7	< 0.1
Aug '81	22.0	1.98	5.2	2.5	8.8	0.11	5.4	7.7	7	< 0.1
Sep '81	21.0	1.85	4.8	2.4	8.0	0.09	5.8	7.7	8	< 0.1
Oct '81	19.0	1.74	4.5	2.3	7.7	0.09	6.2	6.8	8	< 0.1
Nov '81	19.0	1.65	4.3	2.5	9.2	0.12	6.6	6.7	16	0.7
Dec '81	17.0	1.81	4.7	2.7	10.8	0.16	5.9	4.5	17	0.5
Design	17.5	3.70	9.6	6.3	25.0	0.18	2.9	8.0	20	15.0

NOTES:

1. Mixed Liquor, Volatile Suspended Solids (MLVSS) was approximately 70% of Mixed Liquor Suspended Solids (MLSS).
2. RBC Effluent $SBOD_5$ is approximately 35 to 40% of RBC Effluent Total BOD_5.
3. Mean Cell Residence Time equals Lbs MLSS divided by Lbs Solids wasted (as waste activated sludge and secondary effluent TSS).

CONCLUSIONS

The design method for RBC sizing proposed by Antonie(1) and as presented on Figure 3 is a valid design method for sizing a RBC process as a biological roughing process ahead of an activated sludge process.

Using a RBC process as a biological roughing process ahead of an activated sludge process is a workable and cost-effective method for upgrading an existing activated sludge process.

Periodic sludge withdrawal from RBC tanks may help prevent white patchy growth on the RBC bio-film.

Sludge handling/treatment recycle streams adversely affect RBC performance.

The RBC biological roughing process enhances overall plant performance by producing biological solids which encourage nitrification within the activated sludge process and aid in secondary effluent clarification.

REFERENCES

1. Antonie, R. L., "Rotating Biological Contactor for Secondary Wastewater Treatment," presented at the Culp/Wesner/Culp WWT Seminar on October 27-28, 1976, held at South Lake Tahoe, Stateline, Nevada.

2. Autotrol Wastewater Treatment Systems Design Manual, dated 1979, by Autotrol Corporation Bio-systems Division, Milwaukee, Wisconsin.

USE OF SUPPLEMENTAL AERATION AND PH ADJUSTMENT TO IMPROVE NITRIFICATION IN A FULL SCALE ROTATING BIOLOGICAL CONTACTOR SYSTEM

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INTRODUCTION

Nitrification was substantially improved following the installation of supplemental aeration in a full scale rotating biological contactor (RBC) system designed to treat domestic wastewater to an effluent level of 10 mg/L of 5-day total biochemical oxygen demand (BOD_5) and 2 mg/L ammonia-nitrogen (NH_3-N). A slight improvement in nitrification was also observed when the pH of the wastewater was adjusted from 6.6 to 8.4 with soda ash for an extended period. This RBC system is the biological treatment portion of a 6 MGD wastewater treatment plant (WWTP) serving a major U.S Army installation with an effective population of 40,000. The RBC units are arranged in 6 treatment banks of 6 stages each with the first 3 stages in each bank intended for BOD_5 removal and the last 3 stages for NH_3-N removal. Performance of the RBC system was evaluated during summer conditions (wastewater temperature of 26° C and system flow of 4.5 MGD) before and after the addition of 8 cfm of supplemental aeration per lineal foot of RBC shaft in the first 2 stages of each bank. The effect of pH adjustment was evaluated by comparing a control bank to a parallel bank in which up to 1200 pounds per day of soda ash was added to the third stage for 7 weeks.

" The opinions or assertions contained herein are the private views of the author and are not to be construed as official or as reflecting the views of the Department of the Army or the Department of Defense."

Prior to the aeration of the wastewater, dissolved oxygen (DO) limiting conditions (1.0 mg/L or less) existed in the first 4 stages, the white sulfur bacteria, Beggiatoa, predominated on the media, 75% of the BOD_5-S (soluble) removal occurred in the first 3 stages, and 59% of the applied NH_3-N was oxidized. With the supplemental aeration, DO was never less than 1.5 mg/L, Beggiatoa was sparsely present on only the first stage media, 96% of the BOD_5-S removal occurred in the first 3 stages, and 86% of the NH_3-N was oxidized yielding an effluent concentration of 2.1 mg/L. The design nitrification rate of 0.3 pounds of NH_3-N removed per 1000 ft^2 per day existed only when the DO level was above 2.5 mg/L. Although adjustment of pH produced questionable results, an 11% improvement in NH_3-N removal was briefly observed as compared to the control bank with just supplemental aeration.

Conclusions are that NH_3-N removal is dependent on prior BOD_5-S removal so that there is not competition between BOD_5 and NH_3-N removal organisms for space in the latter stages. Low DO spread BOD_5-S removal into stages where nitrification was to occur. More importantly, the nitrification rate was limited by low DO levels. The need to provide at least 2.5 mg/L of DO in the stages where maximum nitrification is expected was clearly shown. Some benefit may be gained through operation in higher pH ranges; however, the design nitrification rate was achieved in the 6.6 pH range.

BACKGROUND

Treatment Plant

The WWTP was upgraded in 1977 from a trickling filter system to a RBC system in order to provide both secondary treatment and nitrification. The upgraded plant was designed to meet the National Pollutant Discharge Elimination System (NPDES) permit limitations shown in Table 1. A schematic diagram of the unit processes is shown in Figure 1. Design capacity of the WWTP is 6 MGD, while maximum hydraulic capacity is 18 MGD.

The RBC system, shown in Figure 2, consists of 36 mechanically driven RBC units arranged in a matrix of 6 treatment banks, each with 6 stages. The first 3 stages

TABLE 1. NPDES Permit Parameters and Limitations

Parameter	Monthly Average, Summer (May 1 through October 31)
pH	6.0 - 9.0
Chlorine Residual	Min conc to comply with FC limit
Fecal Coliform (FC)	200/100 mL
Suspend Solids (SS)	30 mg/L
Five-day Biochemical Oxygen Demand (BOD_5)	10 mg/L
Ammonia Nitrogen (NH_3-N)	2.0 mg/L
Dissolved Oxygen (DO)	>6.0 mg/L

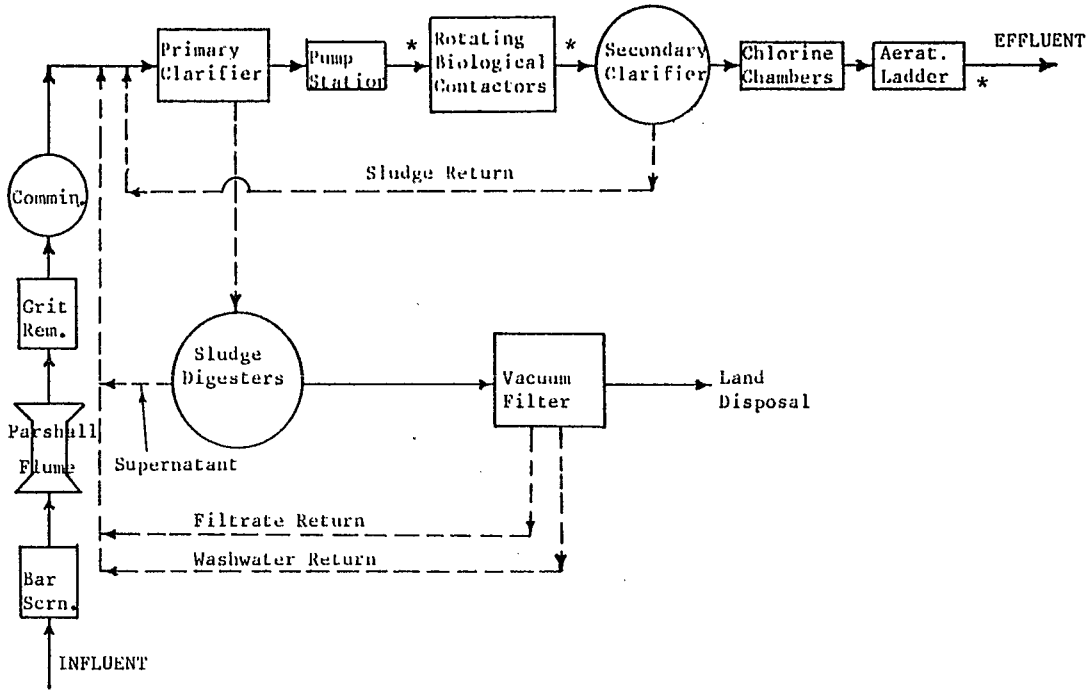

Figure 1. Wastewater Treatment Plant Flow Diagram. (* denotes composite sample point location)

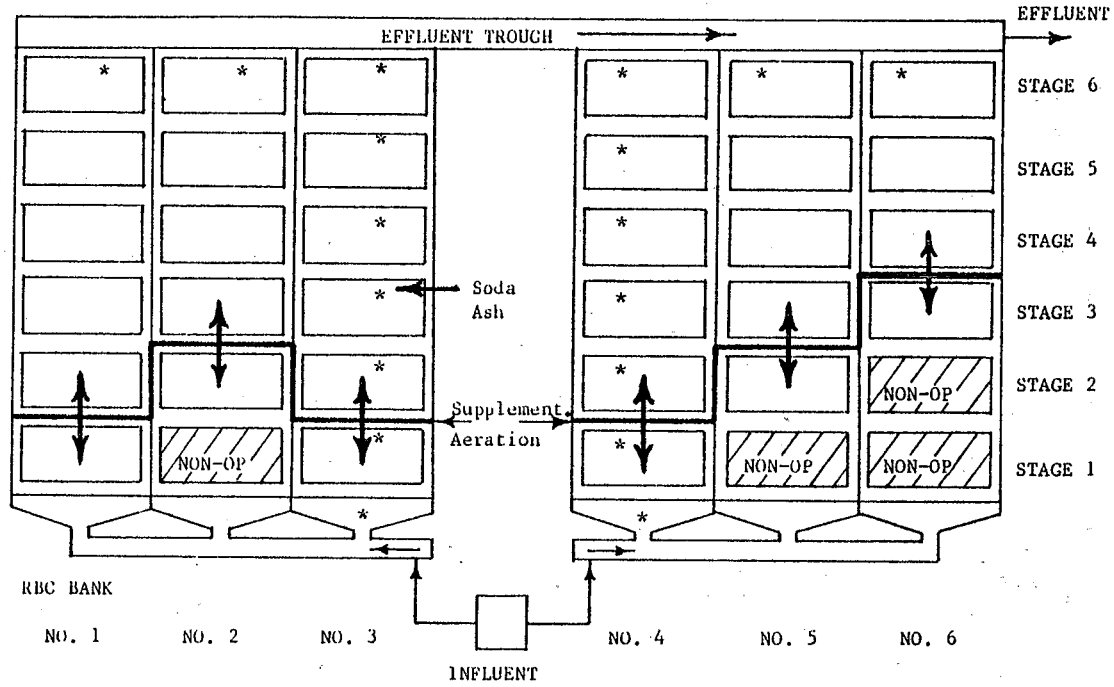

Figure 2. Rotating Biological Contactor System, (* denotes grab sample point location)
(NON-OP indicates stage was non-operational due to shaft failure)

in each bank have regular density media (100,000 ft^2) for BOD_5 removal; the last 3 have high density (150,000 ft^2) for NH_3-N removal. The RBC shaft with media measures 25 ft long and 12 ft in diameter. Each RBC is positioned in a concrete tank with an approximate volume of 16,500 gal. The stages are separated by underflow baffles which provide plug flow through the bank. Based on a dye study (1), the hydraulic detention time across 6 stages is 2 hours and 30 minutes at a flow rate of 5.5 MGD.

The BOD_5 removal part of the RBC system was designed based on hydraulic loading (gpd/ft^2) versus BOD_5 removal (per cent) curves using an influent BOD_5 concentration of 140 mg/L. The overall hydraulic loading is 1.33 gpd/ft^2 at the 6 MGD design flow. Specific removal rates were used to size the NH_3-N removal part of the RBC system. The design was based on an influent NH_3-N concentration to stage 4 of 15.8 mg/L. A removal rate of 0.28 pounds NH_3-N removed per 1000 ft^2 of media surface per day was used for NH_3-N removal down to 5 mg/L. Removal from 5 to 2 mg/L is to be done at 0.20 pounds NH_3-N removed per 1000 ft^2 per day(1).

Previous Studies

Summer and winter studies (August 1978 and January 1979) were conducted by Hitdlebaugh and Miller (1, 2, 3) to evaluate the performance of the upgraded WWTP. They found that the RBC system performed at less than design expectations for BOD_5 and NH_3-N removal. This was attributed both to DO limiting conditions (less than 1 mg/L) in several RBC stages and to relatively low pH (less than 7.0) in the latter RBC stages. During the winter study when DO limiting conditions did not exist, the RBC system actually removed more NH_3-N than during the summer study in spite of the winter wastewater temperature of 13°C. Analyses of samples for BOD_5-S and nitrification-suppressed BOD_5 was found to be essential for the evaluation of WWTP's designed for nitrification. Recommendations for future RBC system designs called for the use of supplemental aeration to overcome limiting DO levels and chemical feed to maintain optimum pH levels.

Supplemental Aeration

A supplemental aeration system was installed in February 1981 to improve BOD_5 and NH_3-N removal in the RBC system. An equally important benefit was expected to be the physical stripping of excess attached biological growth, thereby reducing the possibility of further RBC shaft failures. Between May 1980 and August 1981, 4 stages had become non-operational due to shaft failure. Diffusers were installed in the first 2 working stages of each bank as shown in Figure 2. The circular coarse bubble diffusers (4 per shaft) are offset from the shaft center plumb line by about 2 ft. Clearance between the RBC media and tank bottom ranges from 6 in. at stage 1 to 15 in. at stage 6. Air is provided by 2 blowers, each with 1200 cfm capacity. Shaft weight measurement devices called "load cells" were also installed at this time on bank 4. Using these load cells, the operators can make adjustments to the air flow rate to insure that the maximum shaft weight specified by the manufacturer is not exceeded.

METHODOLOGY

Objectives and Materials

The objectives of this study were to evaluate the effect of the supplemental aeration on the RBC system performance and to assess the potential benefit from pH adjustment by chemical addition to levels considered optimum for nitrification. Since the NH_3-N discharge limit had always been exceeded in the month of August (highest wastewater temperature and lowest DO) and since the August 1978 study by Hitdlebaugh and Miller provided an excellent baseline for conditions existing prior to the use of supplemental aeration, August 19-25, 1981 was selected for the study period. RBC bank 4 (see Figure 2) had suffered no structural damage and continued, as in the previous study, to be used as the primary bank to evaluate the internal performance of the RBC system. Because bank 3most resembles bank 4, it was used as the experimental pH adjustment bank. The decision to add soda ash at the end of stage 3 (see Figure 2) was based primarily on pilot scale RBC studies done by Stratta and Long(4). Their studies indicated that pH adjustment with soda ash yielded NH_3-N removal as good as with lime and did not cause solids precipitation problems. Their studies also showed that the nitrifying organisms need 5 weeks to acclimate to a higher pH.

A chemical feed system consisting of a 500 gal. tank with flash mixer and a 30 gpm capacity centrifugal pump was operated for 7 weeks before the August sampling period. Soda ash solutions were made 4 times per day using either 300 lbs. (7.2 % solution) or 400 lbs. (9.6 % solution) of dry soda ash. The solution was pumped at a constant 1 gpm rate to 2 points ahead of the effluent baffle in stage 3. The higher concentration solution was used during the times of the day when peak wastewater flows occurred. An automatic pH control system, later seen to be essential, was outside the scope of this study. Despite equipment problems and washout by rain induced high flows, the pH in stage 3 was maintained between 8.1 and 9.3 for 4 of the 7 weeks.

Sampling and Analyses

The sampling and analytical program from the 1978 study was duplicated as closely as possible to permit accurate comparison of results from both studies. Twenty-four hour flow proportioned composite samples were collected for 7 days at the RBC system influent and effluent, and at the WWTP effluent. Grab samples of the RBC bank influent and wastewater in each of the 6 stages of banks 3 and 4 were collected at 5 times during the study to determine changes in wastewater characteristics through the RBC system. Sampling times were selected to correspond to those used in the previous study. Grab samples of the wastewater in stage 6 of the other 4 banks were also collected. Temperature, DO, and pH data were taken during each sample period using portable instruments. Sample point locations are shown on Figures 1, 2, and 2a.

All sample analyses were conducted onsite by the Environmental Chemistry Division of the U.S. Army Environmental Hygiene Agency. A mobile laboratory was set up at the WWTP for both studies. Nitrification was suppressed using the ammonium chloride method (5) in order to determine the relative BOD_5 exerted by carbonaceous and nitrogenous substances. Tests for soluble BOD_5 and TOC were conducted on the filtrate passing through a 0.45 micron filter. All sampling and analyses were conducted in accordance with "Standard Methods for the Examination of Water and Wastewater" (6) or "Methods for Chemical Analysis of Water and Wastes" (7).

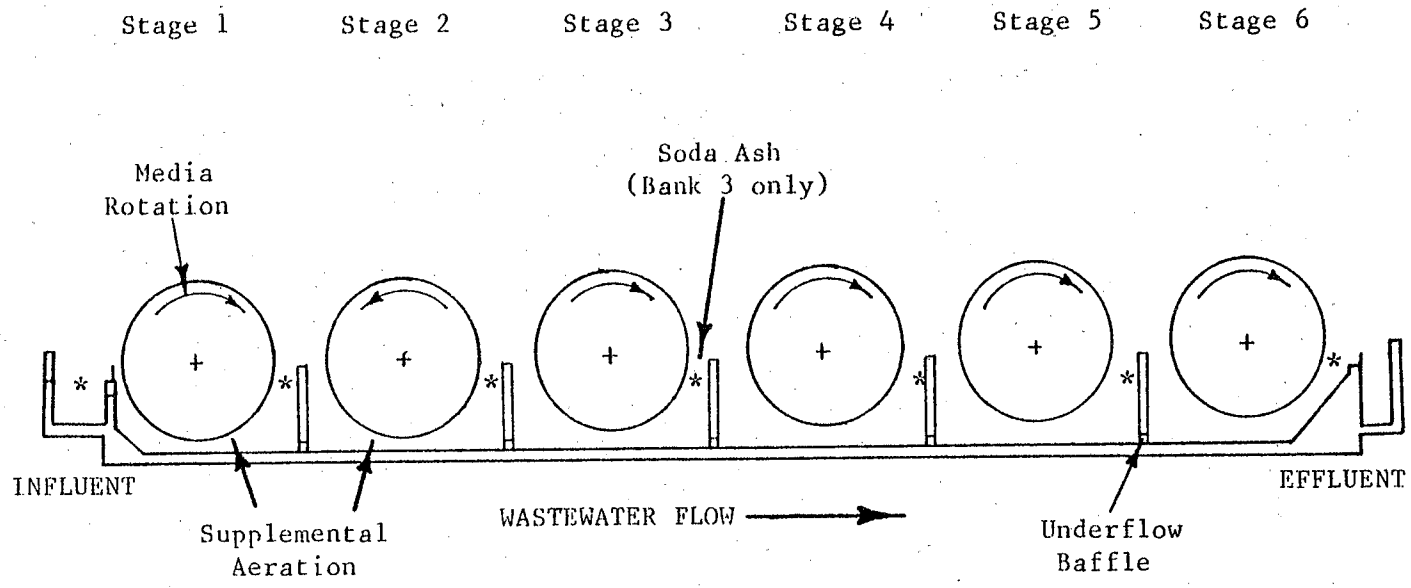

Figure 2a. Rotating Biological Contactor Sample Point Locations- Banks 3 and 4.
(* denotes grab sample point location)

FINDINGS

The sulfur oxidizing bacteria, Beggiatoa, which had been abundant during low DO conditions, was sparsely present on only the first stage media under the new aerated conditions. The biomass growth was much thinner and more uniform in the aerated stages, possibly due to physical stripping action of the air.

Hydraulic and organic loading rates are shown in Table 2. Wastewater characteristics (from composite samples), in and out of the RBC system, and at the WWTP effluent are shown in Tables 3 and 4. The performance (from grab samples) of individual banks is shown in Table 5. Changes in wastewater characteristics (from grab samples) as it passes through each RBC stage are shown in Tables 6-8 and Figures 3 and 4. BOD_5-S and NH_3-N removal rates in each stage are shown in Figures 5 and 6.

Loading and operating conditions were essentially the same for both studies (Tables 2 and 3). The RBC system effluent NH_3-N concentrations were improved from 6.2 mg/L in 1978 to 2.1 mg/L in 1981 because DO limiting conditions were eliminated by the supplemental aeration (Tables 3 and 6, Figure 3). Had all 36 RBC stages been operational, RBC system effluent NH_3-N concentrations would have been in the 1.3 to 1.8 mg/L range achieved by indivdual banks 1 and 4 (Table 5). As predicted by Hitdlebaugh and Miller (2,3), the impact of the aeration on BOD_5-S removal was not to increase the amount removed (Table 3), but to concentrate removal in the first 2 stages (Table 6, Figure 3) at a higher rate (Figure 5). Because BOD_5-S removal was occurring more efficiently, more space for nitrifying organisms was available in the early stages and nitrification actually began in stage 1 (Table 6, Figure 3) with the peak nitrification rate occurring in stage 3 (Figure 5). Although BOD_5 and NH_3-N removal organisms do compete for space, the major factor limiting nitrification was the low wastewater DO levels (Figure 6). A minimum DO level of 2.0 mg/L available at the location where nitrification is expected to occur has been suggested by others (8). The design nitrification rate (1,9) was observed at the 2.5 mg/L DO level (Figures 5 and 6). The largest DO drop across any of the stages was across the stage with the maximum nitrification rate (Figure 6). This is expected since NH_3-N removal requires 4.6 times as much DO as BOD_5-S removal (8). Mechanical reaeration from turning RBC media does not bring the DO level back up to 2.0 mg/L until 2 stages later (Figure 6); therefore, supplemental aeration provided to stage 3 would improve nitrification rates and overall NH_3-N removal.

The improved NH_3-N removal by the RBC system can be seen in the components of total BOD_5 in the WWTP effluent (Table 4). Before aeration, the nitrification process was continuing out the end of the plant indicated by a total BOD_5 of 11 mg/L and nitrogenous BOD_5 of 7 mg/L (total minus carbonaceous). With aeration, nitrification shifted back up into the RBC system indicated by a total BOD_5 of 5 mg/L and nitrogenous BOD_5 of 2 mg/L.

The effects of pH adjustment were disappointing because it is well documented that higher pH levels (8.0 to 8.5) are optimum for nitrification (4). A side by side comparison of banks 3 and 4, showed no improvement in total NH_3-N removed (Table 7), although a different removal rate pattern was observed. The set of grab samples on August 19 did show an 11% improvement (Table 8); however, this was questionable because NO_2/NO_3-N levels did not support the NH_3-N levels. The major reason for the poor performance was probably the wide pH fluctuations (8.1 to 9.3) inherent in the chemical feed system which did not let the nitrifying organisms acclimate to a constant pH.

Amazingly, the maximum design nitrification rate of 0.3 lbs NH_3-N removed per 1000 ft^2 per day was observed at a PH of 6.6 (Figures 4 and 5). The overall NH_3-N removal of 86% was the same as that found by Stratta and Long (4), but the maximum removal rate was less than half of their observed value. Under design load conditions (6 MGD), pH adjustment may be needed to achieve the desired NH_3-N removal with existing media area.

SUMMARY

The supplemental aeration of the RBC system eliminated nuisance organisms, enhanced BOD_5-S and NH_3-N removal, and provided the operational flexibility to control the thickness of biomass growth on the media. Placement of air diffusers in not only BOD_5-S removal stages, but also in NH_3-N removal stages should be considered. DO levels should be maintained at 2.5 mg/L in stages where maximum nitrification rates are expected.

ACKNOWLEDGEMENT

The author would like to thank Mr. John A. Hitdlebaugh and Major Roy D. Miller for their helpful advice in the design of this study. A special thanks is extended to Mr. Charles I. Noss, Mr. Kenneth A. Bartgis, and Captain Edmund Kobylinski from the US Army Medical Bioengineering Research and Development Laboratory for their efforts in support of the study.

The data shown in the following Tables and Figures from the August 15-21, 1978 study (without supplemental aeration) was extracted from the paper entitled "Full-Scale Rotating Biological Contactor for Secondary Treatment," by John A. Hitdlebaugh and Roy D. Miller, presented at the first National Symposium/Workshop on Rotating Biological Contactor Technology, Champion, PA (1980).

TABLE 2. RBC System Loading Rates*

Location	August 15-21, 1978	August 19-25, 1981	Design**
	Hydraulic Loading (gpd/ft^2)		
Six Stages	1.0	0.93	1.33
First Stage	7.5	7.0	-
	Organic Loading (lbs BOD_5/1000 ft^2/day)		
Six Stages	0.60	0.68	1.56
First Stage	4.5	5.08	11.7
	Organic Loading (lbs BOD_5-S/1000 ft^2/day)		
Six Stages	0.18	0.15	-
First Stage	1.31	1.11	4.0 ***

*Based on data in Table 3.

**Based on design flow of 6.0 MGD, RBC influent concentration of 140 mg/L BOD_5, and RBC stage 4 influent concentration of 15.8 mg/L NH_3-N.

***This level recommended to prevent DO limiting conditions in the first stage (9).

TABLE 3. RBC System Influent and Effluent Characteristics[1]

Parameter[6]	August 15-21, 1978[2] Avg Flow = 4.5 MGD[4] Temp = 26°C		August 19-25, 1981[3] Avg Flow = 4.2 MGD[5] Temp = 25°C	
	Influent	Effluent	Influent	Effluent
Conductivity (μmho/cm)	960	930	890	855
Total Alkalinity	158	90	154	84
SS	69	63	55	41
BOD_5	72	61	87	34
BOD_5-soluble	21	4	19	4
TOC[5]	42	24	55	27
TOC-soluble	23	11	21	11
TKN	21	8.9	22	6.8
NH_3-N	16.0	6.2	16.0	2.1
NO_2/NO_3-N	0.05	8.9	<.01	13

[1]Values shown are average of 7 - 24 hr composite samples.
[2]Without supplemental aeration.
[3]With supplemental aeration.
[4]Sum of STP influent flow (3.7 MGD) and recirculated flow (.8 MGD).
[5]Sum of STP influent flow (3.9 MGD) and recirculated flow (.25 MGD).
[6]All units are mg/L unless otherwise noted.

TABLE 4. Wastewater Treatment Plant Effluent Values[1]

Parameter[2]	August 15-21, 1978[3]	August 19-25, 1981[4]
pH (Standard units)[5]	6.7	6.9
BOD_5 - total	11	5
BOD_5 - soluble	2	2
BOD_5 - carbonaceous	4	3
NH_3- N	6.2	2.6
SS	9	6.4
Flow (MGD)	3.7	3.9

[1]Values shown are average of 7 - 24 hr composite samples.
[2]All units are mg/L unless otherwise noted.
[3]Without supplemental aeration.
[4]With supplemental aeration.
[5]Average of 7 daily measurements made at various times with a portable instrument.

TABLE 5. RBC System Performance by Bank[1]

August 19-25, 1981; Supplemental Aeration in all Banks; pH Adjustment in Bank No. 3
Wastewater Temp = 25°C; Total System Flow = 4.2 MGD

	Stages With Shaft Failure	pH	T. Alk	DO	BOD_5-S	TKN	NH_3-N	NO_2/NO_3-N
RBC Influent	-	6.9	158	2.0	29	22	18	<0.01
Bank 1 Effluent	-	6.4	53	3.0	4	5.0	1.3	14
2	No. 1	6.4	67	2.1	4	6.7	2.9	12
3	-	7.0	143	3.1	3	5.9	2.1	13
4	-	6.3	57	2.7	4	5.8	1.8	13
5	No. 1	6.3	71	2.1	4	9.9	3.7	11
6	No. 1 & 2	6.4	87	2.1	5	10	4.9	8.9

[1]Average of results from 5 groups of grab samples collected at the same set of times for each study. All units are mg/L except for standard pH units.

TABLE 6. RBC Bank 4 Performance by Stage[1]

August 15-21, 1978; Bank 4; Without Supplemental Aeration
Wastewater Temp = 26°C; Flow = .75 MGD

	pH	T. Alk	DO	BOD_5-S	TKN	NH_3-N	NO_2/NO_3-N
RBC Influent	6.8	159	3.4	24	23	16	<0.04
Stage 1	6.8	159	1.4	15	22	16	<0.04
2	6.8	159	0.7	9	20	16	0.2
3	6.75	149	0.8	6	19	15	1.4
4	6.6	126	1.3	4	14	12	4.4
5	6.55	104	1.9	3	11	8.3	7.7
6	6.5	93	2.2	3	9	6.6	9.7

August 19-25, 1981; Bank 4; With Supplemental Aeration
Wastewater Temp = 25°C; Flow = .70 MGD

RBC Influent	6.8	158	2.3	29	22	18	<0.01
Stage 1	6.8	155	1.7	12	21	17	<0.01
2	6.8	142	2.7	7	19	15	1.2
3	6.6	115	1.5	5	14	10	4.7
4	6.4	86	1.6	5	8.9	6	8.6
5	6.3	67	2.1	4	7.5	3	12
6	6.3	57	2.7	4	5.8	1.8	13

[1]Average of results from 5 groups of grab samples collected at the same set of times for each study. All units are mg/L except for standard pH units.

TABLE 7. Effects of pH Adjustment - Average of 5 Grab Samples[1]

August 19-25, 1981, Wastewater Temp = 25°C; Flow per Bank = .70 MGD
Bank 4; Supplemental Aeration Only

	pH	T. Alk	DO	BOD_5-S	TKN	NH_3-N	NO_2/NO_3-N
RBC Influent	6.8	158	2.3	29	22	18	<0.01
Stage 1	6.8	155	1.7	12	21	17	<0.01
2	6.8	142	2.7	7	19	15	1.2
3	6.6	115	1.5	5	14	10	4.7
4	6.4	86	1.6	5	8.9	6	8.5
5	6.3	67	2.1	4	7.5	3	12
6	6.3	57	2.7	4	5.8	1.8	13

Bank 3; Supplemental Aeration and pH Adjustment in the End of Stage 3.

RBC Influent	6.9	157	1.6	29	22	18	<0.01
Stage 1	6.9	155	1.8	14	21	18	0.02
2	6.9	146	2.4	5	19	14	1.1
3	7.6	186	1.4	6	15	10	4.3
4	7.3	171	1.6	3	10	5	7.7
5	7.0	150	2.0	3	7.2	3	11
6	7.0	143	3.1	3	5.9	2.1	13

[1]Average of results from 5 groups of grab samples collected at the same set of times. All units are mg/L except for standard pH units.

TABLE 8. Effects of pH Adjustment - Best of 5 Grab Samples[1]

August 19, 1981; Wastewater Temp = 25°C; Flow per Bank = .73 MGD

	pH	T. Alk	DO	BOD5-S	TKN	NH_3-N	NO_2/NO_3-N
Bank 4; Supplemental Aeration Only							
RBC Influent	6.8	157	2.7	24	23	19	<0.01
Stage 1	6.8	156	2.2	13	22	18	0.1
2	6.9	144	4.0	6	19	16	1.6
3	6.6	116	1.8	6	15	10	3.9
4	6.3	84	1.6	5	11	6.0	9.6
5	6.2	54	1.8	4	8.4	3.0	13
6	6.3	56	2.5	2	5.6	2.0	14
Bank 3; Supplemental Aeration and pH Adjustment in the End of Stage 3.							
RBC Influent	6.9	159	2.2	18	23	20	<0.01
Stage 1	7.0	158	2.2	12	22	19	0.06
2	7.0	144	2.9	-	20	15	1.3
3	8.4	229	1.6	7	16	10	4.9
4	8.0	193	1.5	6	10	6.0	9.0
5	7.3	153	1.7	4	7.1	3.0	12
6	7.1	139	2.7	4	5.0	1.2	13

[1]Values shown are from the grab samples which demonstrated the best NH_3-N removal of 5 grab samples taken from August 19-25, 1981. All units are mg/L except for standard pH units.

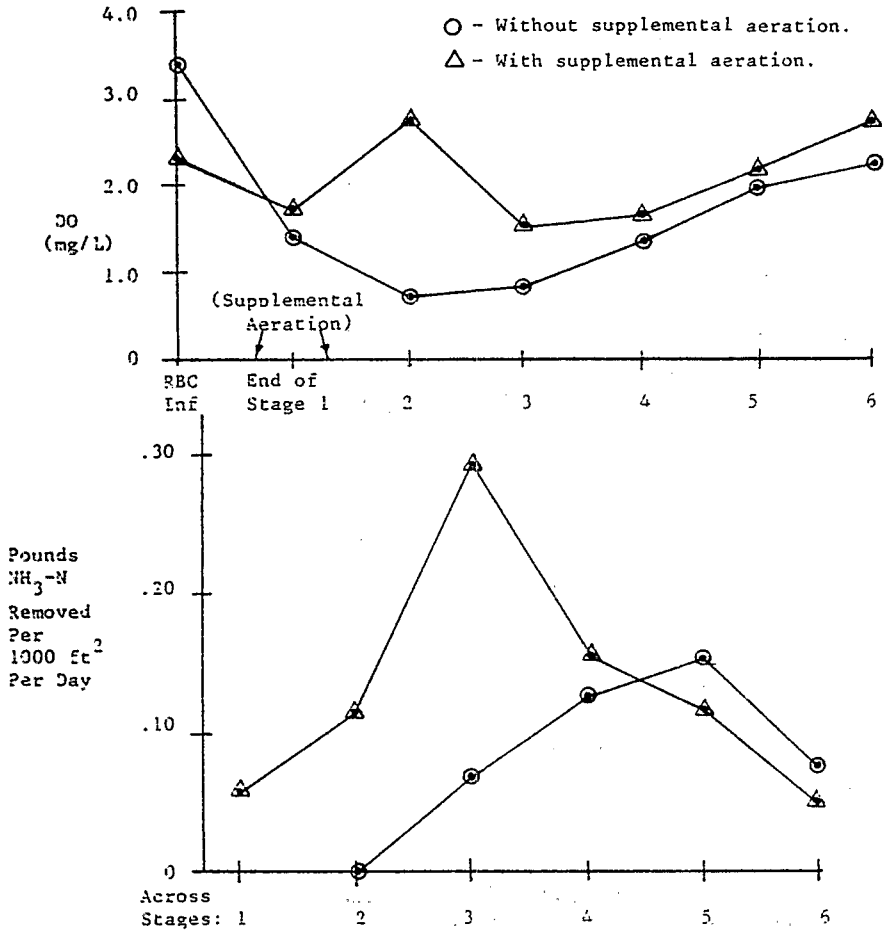

Figure 6. Effects of Dissolved Oxygen Concentration on the NH_3-N Removal Rate (Data from Table 6).

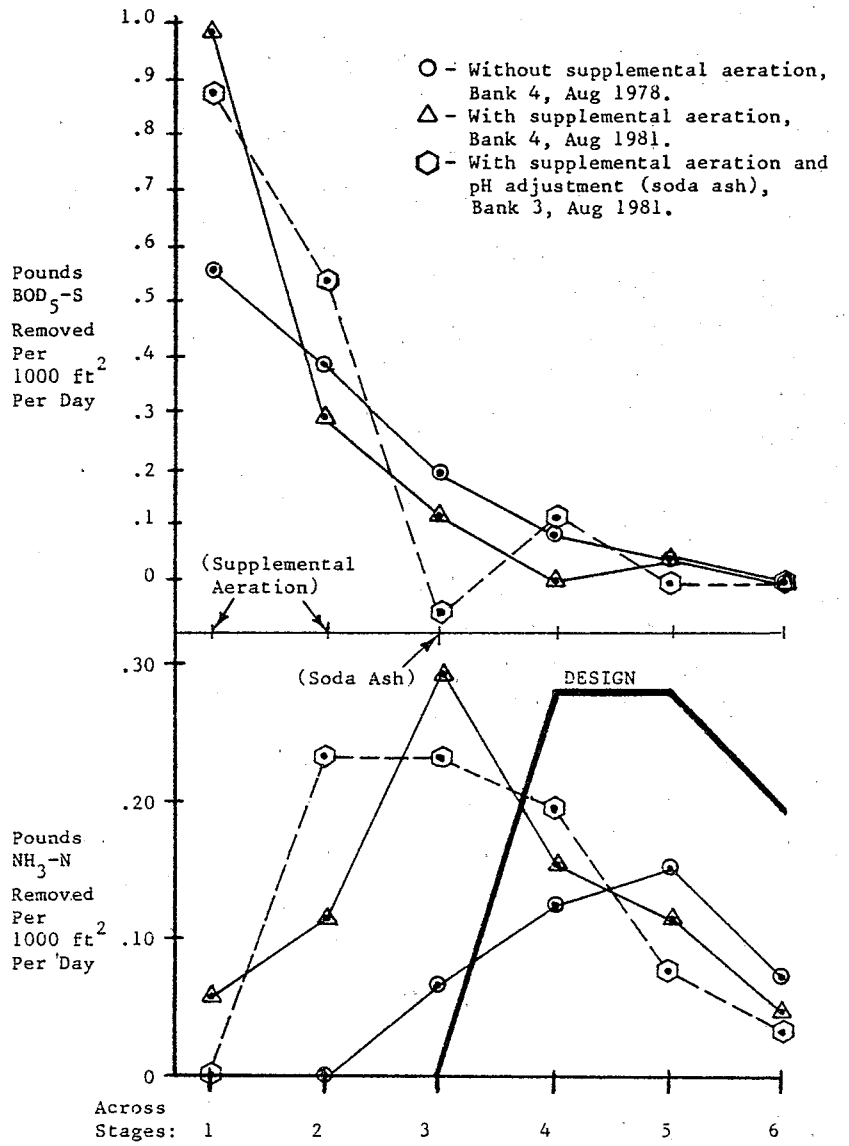

Figure 5. Comparison of Removal Rates With and Without Supplemental Aeration and pH Adjustment (Based on data in Tables 6 & 7).

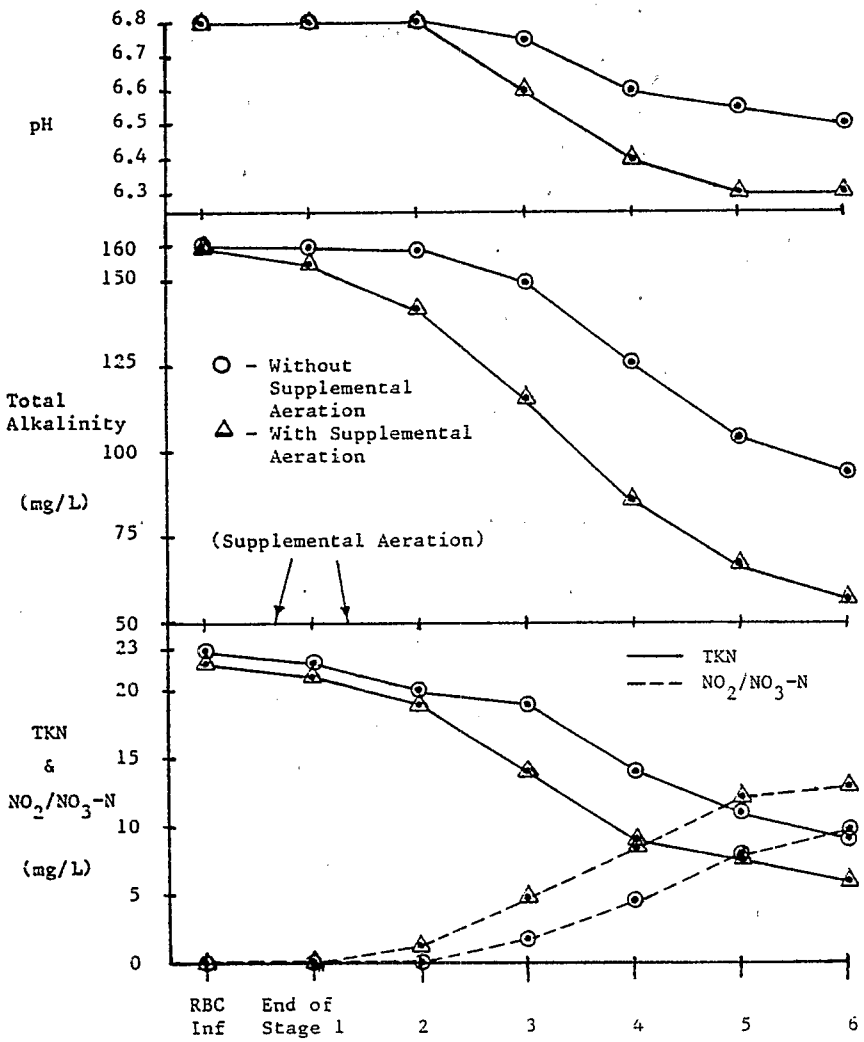

Figure 4. Comparison of pH, Total Alkalinity, TKN, NO_2/NO_3-N Levels Through RBC Bank 4 With and Without Supplemental Aeration (Data from Table 6).

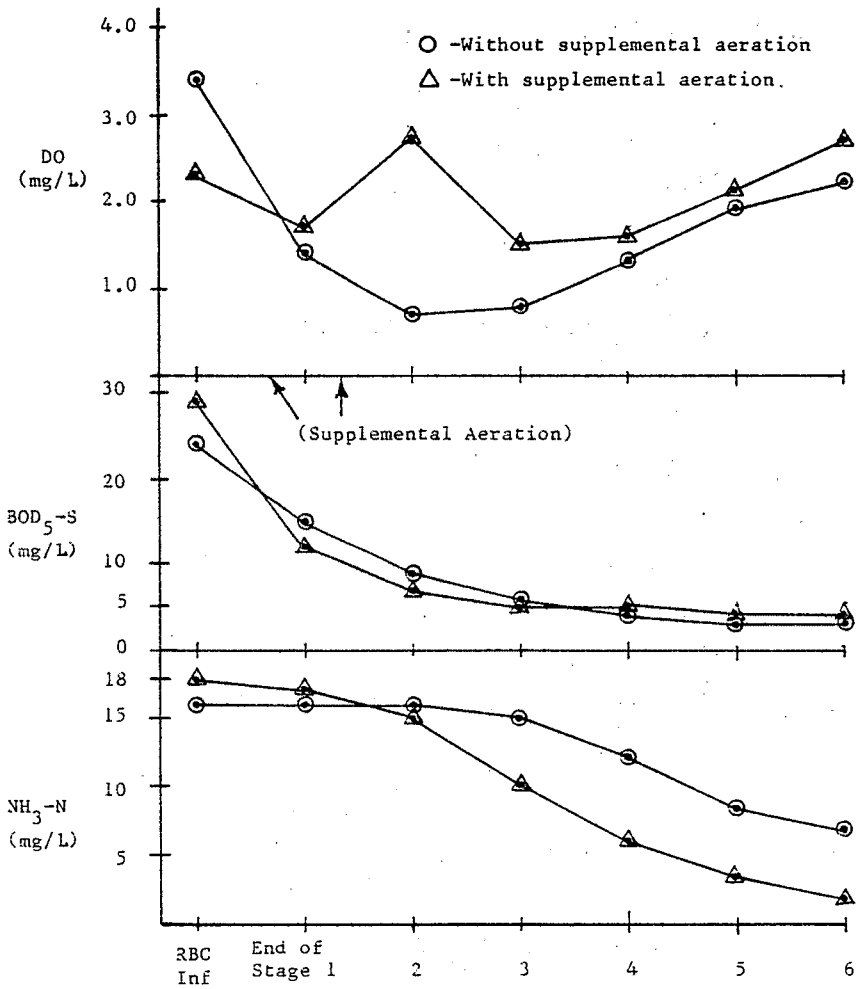

Figure 3. Comparison of DO, BOD_5-S, and NH_3-N Levels Through RBC Bank 4 With and Without Supplemental Aeration (Data from Table 6).

ABBREVIATIONS

BOD_5	5-day total biochemical oxygen demand
BOD_5-S	Soluble BOD_5
$CaCO_3$	Calcium carbonate
cfm	Cubic feet per minute
DO	Dissolved oxygen
Eff	Effluent
FC	Fecal coliform
ft^2	Square feet
gal	Gallon
gpd/ft^2	Gallons per day per square foot
gpm	Gallons per minute
in	Inch
Inf	Influent
lbs	Pounds
mg/L	Milligram per liter
MGD	Million gallons per day
ml	Milliliter
µmho/cm	Micromhos per centimeter
N	Nitrogen
NH_3-N	Ammonia expressed as nitrogen
NO_2/NO_3-N	Nitrite plus nitrate expressed as nitrogen
pH	Negative logarithm of hydrogen ion concentration
SS	Suspended solids
T Alk	Total alkalinity
Temp °C	Temperature in degrees centigrade
TKN	Total Kjeldahl nitrogen
TOC	Total organic carbon
TOC-S	Soluble TOC

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APPLICATION OF ROTARY SCREENS, BIOLOGICAL CONTACTORS, AND GRAVITY PLATE SETTLERS TO TREAT WASTEWATERS IN HOBOKEN AND NORTH BERGEN, NEW JERSEY

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INTRODUCTION

A flow-through compact system composed of rotary screens, biological contactors, and gravity plate settlers was tested by Mayo, Lynch and Associates at a full-scale pilot plant in Hoboken, New Jersey for wastewater treatment. The results of the pilot plant study was applied to the North Bergen Central Sewerage Treatment Plant, North Bergen, New Jersey which was also designed by Mayo, Lynch and Associates. The Utilization of this flow-through system for municipal wastewater treatment has been demonstrated successfully at the North Bergn SPT. This is the first full-scale plant, adopting the concept of flow-through compact system in the world.

The flow-through system is composed of rotoscreen, biological contactors and gravity plate settlers. When wastewater flows into this system, particles greater than 0.02 inch are removed by rotoscreen first. The rotoscreen, which replaces of grit chamber and primary settling tank in conventional treatment system, has the function of primary treatment. The effluent from the rotoscreen then flows through the biological contactors which could be either rotating biological contactors (RBC) or trickling filters. The evaluation of treatment levels at various rates of hydraulic loadings and biological contactor stages was also conducted during the course of this study. The results of the tests were used to establish guidelines for the design of the North Bergen Central Sewage Treatment Plant.

Some advantages of a flow-through system consisting of rotoscreens, biological contactor and gravity plate settlers to wastewater treatment include: reduced land requirements, reduced capital costs as well as a reduction in operation and maintenance costs when compared to a conventional activated sludge process facilities; ability to meet secondary treatment requirements; ability to withstand hydraulic and organic surges; compatibility in good settling characteristics of sloughed sludge with the gravity plate settlers; ease of operation and elimination of wind and thermal disturbances. These and other characteristics of the systm were examined during our evaluation of pilot the plant study, the operation of the full scale North Bergen Central Sewage Treatment Plant demonstrated it.

The object of this paper is to present the results of an investigation into the feasibility study of a flow-through system for wastewater treatment as shown by results from the Hoboken Pilot Plant. The operation data of North Bergen Central Sewage Treatment Plant, which was designed and incorporates the results of the Hoboken Pilot Plant study, were collected to support the concept of the flow-through system. Reluctant data concerning both, the pilot plant study and full-scale treatment plant operation, are presented in this paper.

PLANT OPERATION

Hoboken Pilot Plant

The pilot plant was established within the site of the Hoboken Treatment Plant, Hoboken, New Jersey. There were two flow-through treatment configurations in the pilot plant tested in full scale operations to establish treatability. The first configuration consisted of a rotoscreen, two-stage trickling filter and a gravity plate settler. The second configuration was essentially the same except that the two-stage trickling filter was replaced by a rotating biological contactor. The raw wastewater, which flowed to the full-scale pilot plants by a splitter box shown in Fig 1, was pumped from the head end of the treatment plant just ahead of grit chambers. In this way no solids would be removed prior to the rotoscreen. The rotoscreen is a stainless steel drum of which the periphery is covered with a stainless steel mesh with 0.02 inch openings. The unit is so designed that the sewage passes through the screen and the solid particles larger than 0.02 inches become impinged upon the screen. As the drum rotates the solids are scraped off the screen thereby cleaning the screen. The cleaned portion of screen is then rotated until it again comes into contact with the raw sewage.

The sewage passing through the rotoscreen was divided between the trickling filters and the rotating biological contactor. The trickling filter was constructed of steel containing four sections. Each section has eight foot depth packed with B.F. Goodrich vinyl core media. In order to disperse tne sewage evenly across the top of the trickling filter, 16 evenly spaced nozzles were installed. The gravity of flow of sewage ovr the media forms a biological slime which provides the medium for the biological treatment. The biologically treated sewage then flowed into the LAMELLA gravity settler manufactured by the Parkson Corporation. This unit separates the biological solids and other solids from the sewage by means of sedimentation between inclined plates within the unit.

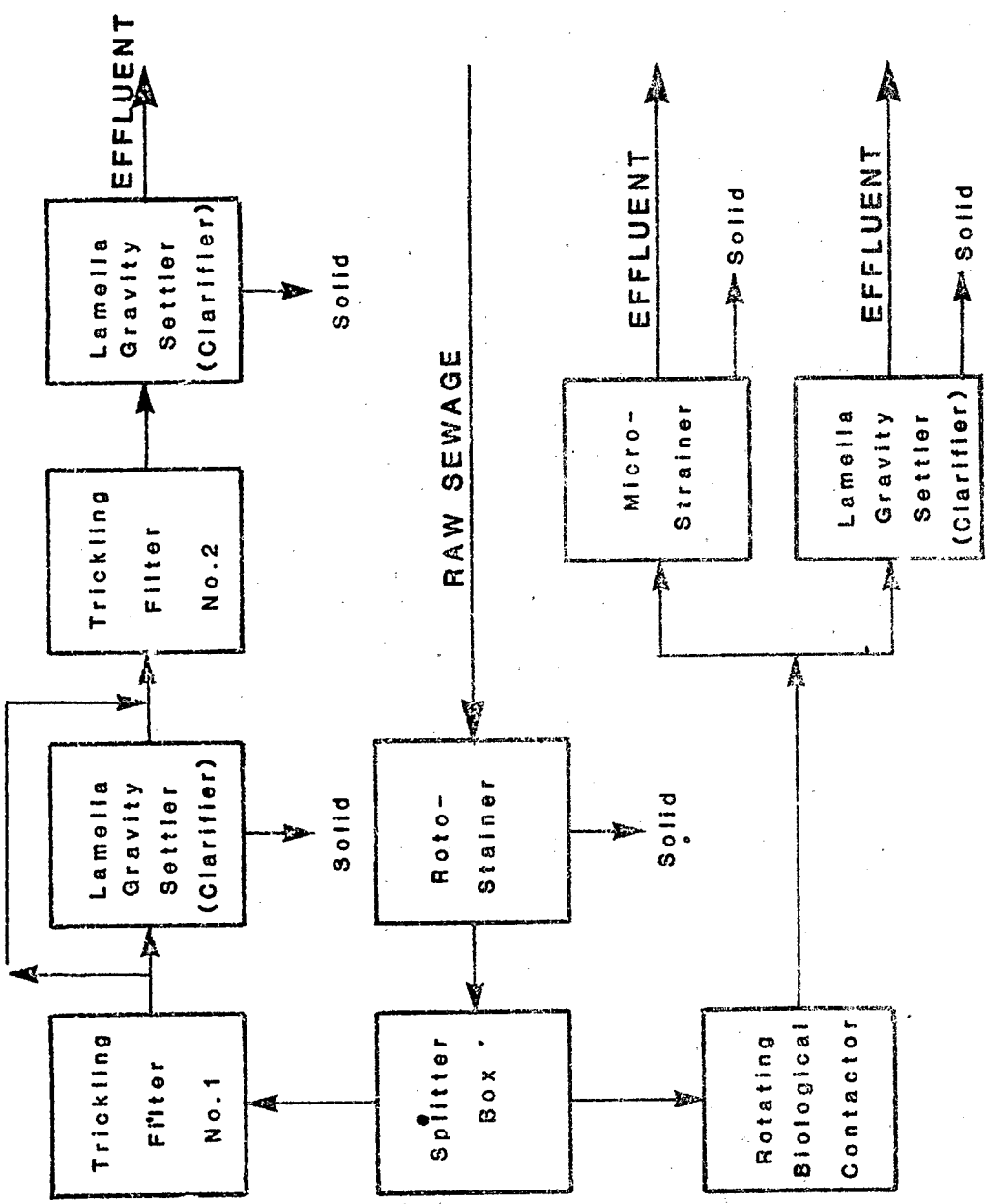

Fig.1 FLOW SHEET FOR PILOT PLANT STUDY

The second means of biological treatment tested was the rotating biological contactor (or bio-disc). The bio-disc consists of circular plastic sheets revolving within a hemi-cylindrical tank, through the sewage flows. The plastic sheets or disks rotate through the sewage growing the biological slime for treatment. The disk used for this pilot plant was 11'-8" long, 12'-10" wide. The unit was divided into five equal stages each containing 8,190 ft^2 of media surface. The circular media sheets were 1'-9" in diameter. The rotating bio-disc unit was manufactured by EPCO-Hormel Co. During the study various stages - two, three, four and five - RBC's were utilized.

The effluent from the rotating bio-disc then flowed by gravity to the microstrainer or LAMELLA gravity settler. The microstrainer was tested to see if the biological solids produced by the rotating bio-disc could be removed or not. The mirco-strainer has a drum covered with a fine mesh having an opening of 35 microns. The microstrainer was manufactured by ZURN.

The rotating bio-disc, microstrainer and electrical controls were contained within a 24' by 16' by 16' wood building to protect them from the elements.

24 hour composite samples were analyzed for BOD_5, dissolved oxygen, settlable solids and suspended solids. Temperature and pH were taken from plant records and grab samples. Those analyses were performed in accordance with Standard Methods (2).

North Bergen Central Sewage Treatment Plant

The Central Sewage Treatment Plant in North Bergen, New Jersey is an application of rotary screens, biological contactors, and gravity plate settlers as a process for treatment of wastewater.

The treatment plant consists of the following major treatment units: bar screens, lift station, self cleaning fine rotary screen for primary grit and suspended solids reduction; rotating bio-disc system for biological treatment; gravity plate settlers for separation of the biomass and suspended solids from the rotating bio-disc system effluent; post aeration to increase dissolved oxygen concentration in the stream; and chlorination system for the final effluent disinfection. This whole process is shown in Fig. 2.

The plant uses a rotary self-cleaning screen in place of the conventional grit chamber and primary clarifiers. The Hydrocyclonics Corporation's Model RSA-36120 with 0.02" opening was selected for design because it will pass more flow and give an effluent with almost identical levels of suspended solids.

The rotating bio-disc system was designed to lower the soluble BOD to 10 mg/l. A hydraulic application rate of 2.4 gpd/ft^2. is expected to accomplish this, using a two stage arrangement to assure that the system be kept aerobic. At the design flow rate, a minimum of 4.16 million square feet of media surface area is expected to provide this degree of treatment. This amount of surface area can be provided by using 32 units arranged as 16 flow streams of two stages in which the area of the first stage is 104,000 ft^2 per shaft and the area of the second stage is 156,000 ft^2 per shaft. Each shaft of discs is installed in a steel tank with each such unit serving as a stage of treatment. The shafts will be rotated at a speed of 1.6 rpm.

The rotating bio-disc system installed in the North Bergen Central Treatment Plant can be summarized as follows:

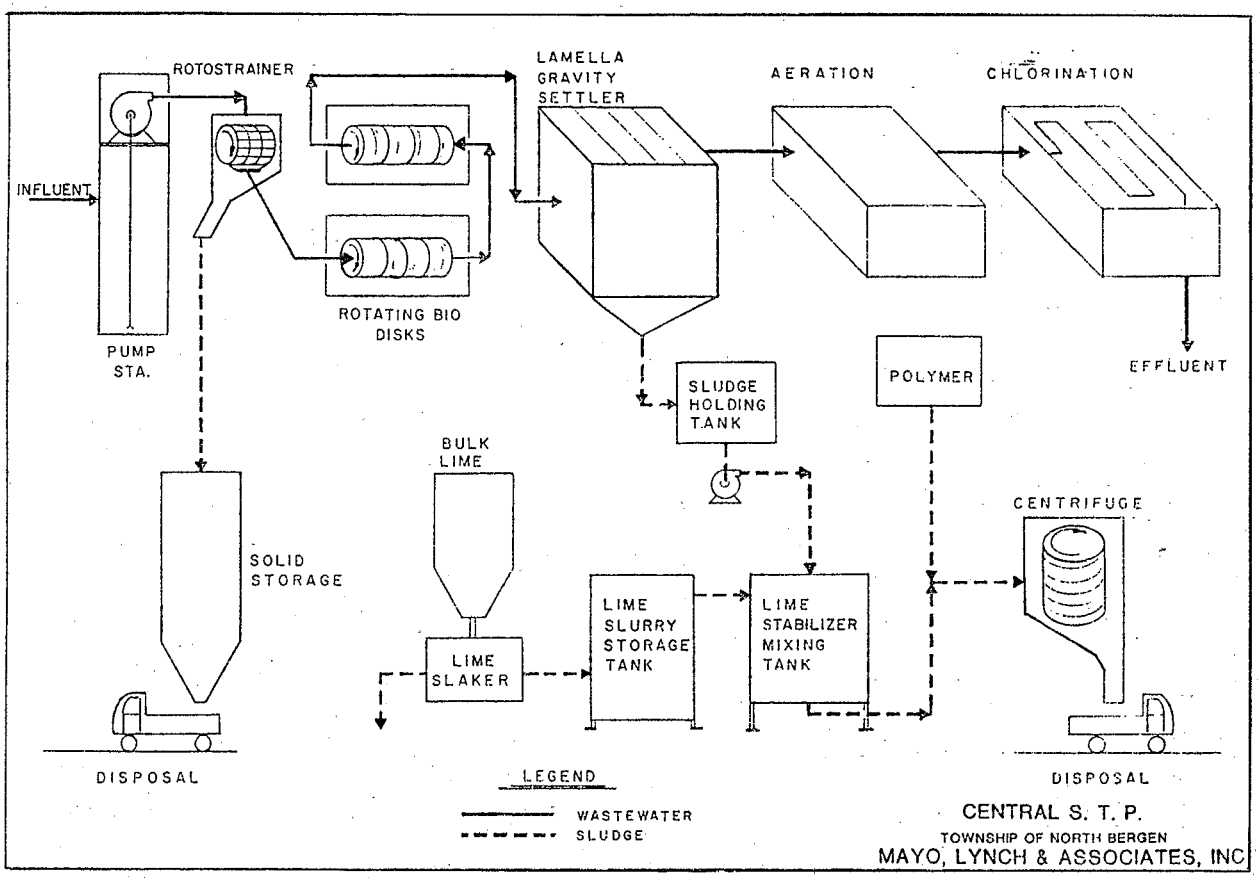

Fig.2 SCHEMATIC NORTH BERGEN CENTRAL SEWAGE TREATMENT PLANT

Type: Autotrol Model 601-251
and 661-251 Total Number of Unit 32 Surface
area/each Unit:
1st Stage 104,000 sq ft. and
2nd Stage 156,000 sq. ft.
Drive Speed: 1.6 rpm
Diameter of Disc. 3.6 meter bio-surf HD shaft
Hydraulic loading: rate 2.4 gpd/ft^2

The Claripak gravity plate settlers installed in the North Bergen Central Treatment plant are summarized as below:

Type	Peabody Welles Series 3000A
Total Number of Unit	7
Dimensions of FRP Sheets	24" wide x 120" long
Inclination	55°
Projected Settling area/each Unit	2,500 ft^2 ea.
Hydraulic Loading	571 gpd/ft^2

The gravity plate settler is suitable for final solids separation following either a rotating bio-disc or trickling filter system which is demonstrated by the pilot plant study in this paper.

The operation of North Bergen Central Sewage Treatment Plant was described according to the parameters of BOD_5, suspended solids, volatile suspended solids, ammonia nitrogen, dissolved oxygen, chlorine residual, fecal coliform, temperature, and pH which was analyzed in the treatment plant by plant personnel.

RESULTS AND DISCUSSIONS

Hoboken Pilot Plant Study

The pilot plant established in the Hoboken Treatment plant was operated in a full-scale facility. The wastewater flows to the pilot plant varied between 13.7 and 56.8 gpm over the test period. But the hydraulic loading rate on the trickling filters or RBC were controlled at desired values by pumping. The wastewater temperatures ranged from 11° to 26°C. The pH levels remained fairly constant at 6.5 $\pm$0.5.

The BOD concentration of the wastewater in Hoboken, New Jersey plant varied throughout the experimentation period. Figure 3 shows the influent BOD concentration of the raw wastewater on the indicatd months. Most of the raw wastewater BOD concentration ranged from 75 to 175 mg/l. The total suspended solids (TSS) concentration of the raw wastewater is shown in Figure 4. TSS varied from 14 mg/l to 175 mg/l averaging 58 mg/l over the test period. But most of them were still in the range of 20 to 80 mg/l.

The performance of each unit in the pilot plant study is described below.

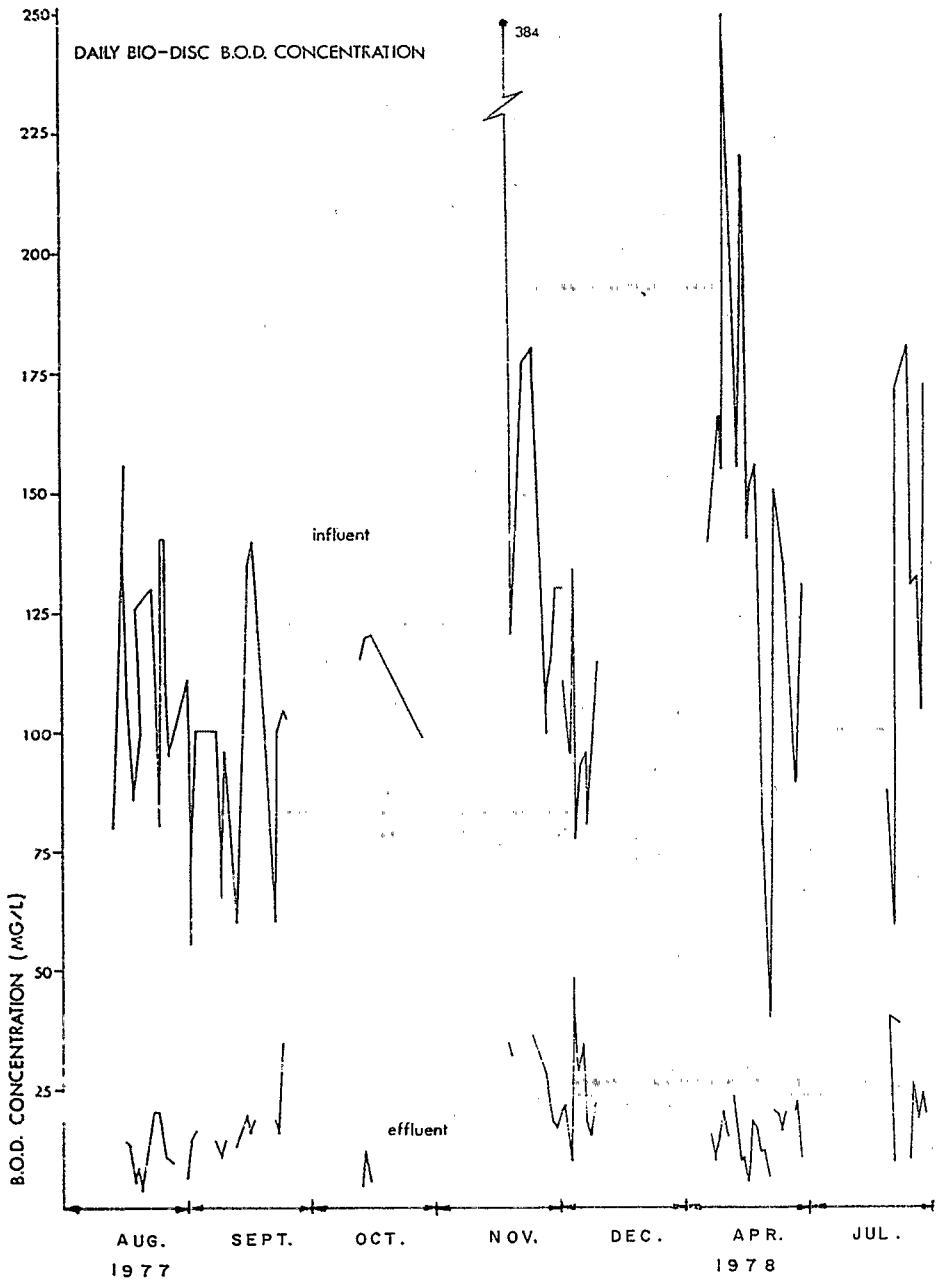

Fig. 3 INFLUENT AND EFFLUENT BOD CONCENTRATIONS BIO-DISC CONFIGURATION

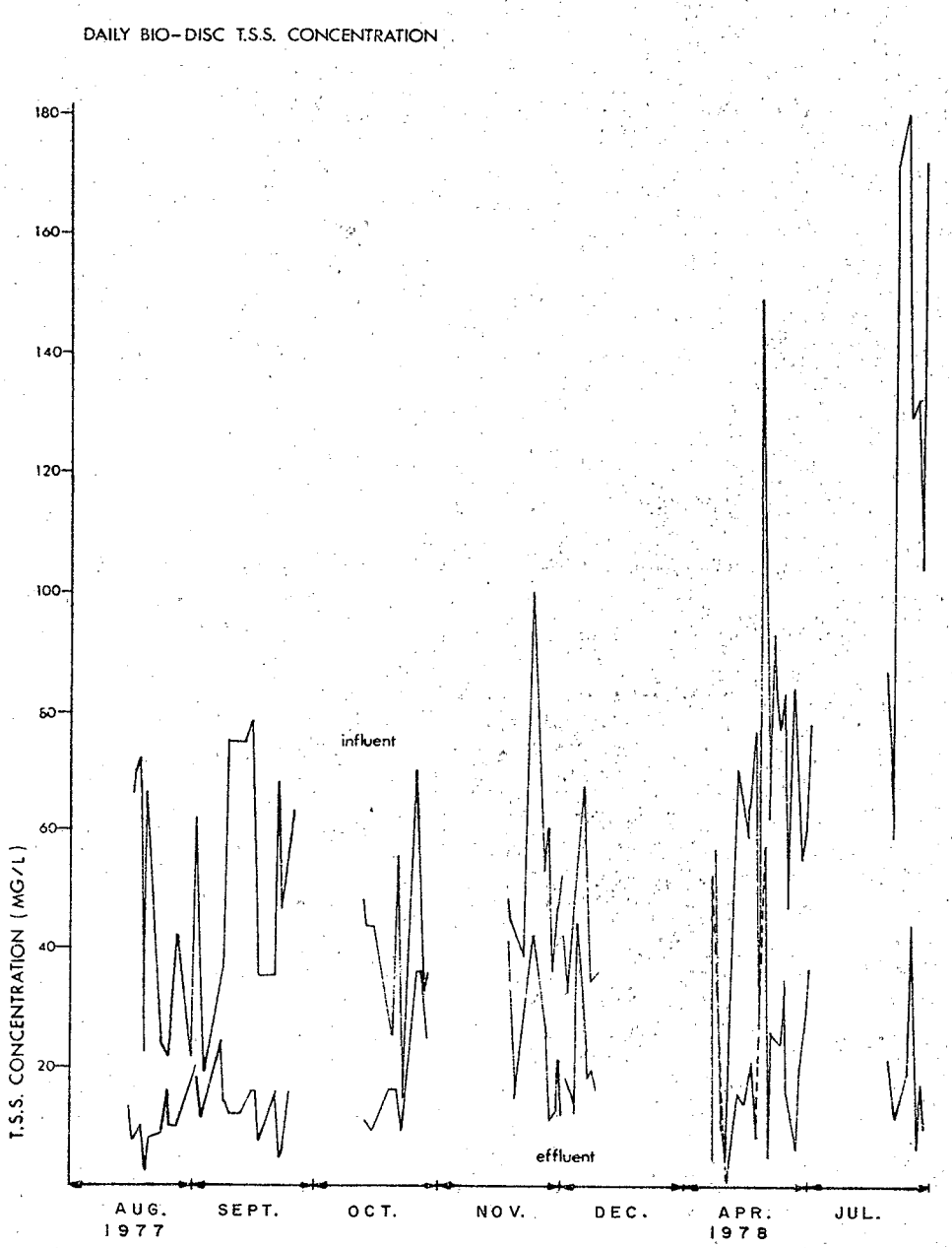

Fig.4 INFLUENT AND EFFLUENT TSS CONCENTRATIONS BIO-DISC CONFIGURATION

Rotoscreen (Rotostrainer)

The openings in the rotoscreen were 0.02 inches against a flow ranging from 13.7 to 56.8 gpm. All solids larger than 0.02 inches were removed. The influent suspended solids concentration ranged from 12 to 162 mg/l, averaging 75 mg/l over the testing period in April, 1978. The effluent readings from the rotoscreen for suspended solids ranged from 5 to 149 mg/l, averaging 65.6 mg/l. The average percentage removal for suspended solids is 7.7%. The actual levels for the influent suspended solids should be somewhat higher than the values indicated in the report. This condition stems from the fact that the sampling devices used to record the data were not capable of accepting gross vegetable and fecal matter found in the raw sewage influent samples. This restriction in the size of particles accepted by the sampler would produce lower than actual results in the influent but would have little effect on the effluent results. Actual suspended solids removal of 15 to 20% have been documented elsewhere (3). It is anticipated that the actual removals were higher than the pilot plant test data indicated. The North Bergen treatment plant design was based on 20% suspended solids removal by the rotoscreen. The actual operations of rotoscreen in the North Bergen Central Sewage Treatment Plant from October, 1981 to January 1982 are shown in Table 9 in which it shows average suspended solids removal was 28.9%.

Compared to the primary settling tank, the rotoscreen can remove less suspended solids. However, it does not significantly affect the performance of trickling filter or rotating bio-disc. The BOD load applied to a trickling filter or rotating bio-disc does not include the fraction which is included in a settleable solids. Therefore, the rotoscreen removes less suspended solids than does a primary settling tank, this will not signficantly increase the BOD loading to the

trickling filter or rotating bio-disc process. All influent settleable BOD is not required to be removed in the primary treatment unit as long as it is not solids which can plug the media and interfere with the biological activity. The settleable solids which are not removed by rotoscreen will be settled in the secondary clarifier.

Trickling Filter

A plastic media trickling filter was used in this study. The effluent from the rotoscreen was fed to the trickling filter. The BOD and SS removals by the trickling filter are shown in Tables 3 and 4. The trickling filter effluent sample was treated to simulate a final clarifier. The composit sample of the trickling filter was settled in a one-liter graduated cylinder for 60 or 30 minutes depending on BOD or SS analyses. It was found during SS testing with the LAMELLA gravity settler that the 30 minute settling test more closely simulated the LAMELLA gravity settler, then the 60 minute settling was used to simulate the final clarifier. Therefore, 30 minute settling was used for SS analyses, and 60 minutes settling for BOD analyses.

As shown in Table 1, at the hydraulic loading rate of 1.4 gpm/ft^2 the overall BOD removals were more than 86%. It can be concluded that the trickling filter provides excellent removals under the application of rotoscreen for the primary treatment. The various BOD concentration of 8.8, 10.0, 18.9, and 16.2 mg/1 from the tower 2 shown in Table 1 exhibits the lower limit with biological treatment. The pilot plant test result has shown that the biological reaction in the trickling filter is similar to the first order kenietic reaction. This implied that the rate of BOD removal will increase as the influent BOD concentration increases. This phenomenon has been demonstrated on wastewater at specific hydraulic loadings for a BOD_5 concentra-

Table 1

Effects of Wastewater Temperature, Intermediate Clarifier and Tower Depth on BOD Removal of Plastic Trickling Filters

pH = 6.5±0.5 and Hydraulic Loading Rate = 1.4 gpm/ft^2

Period	Number of Samples	Average Temperature and Range°C	Screen	Tower 1	Tower 2	Tower 1	Overall	Note
8/15/77 to 8/31/77	12	22 to 26 Ave. 24.5	56 to 157 Ave. 108.3	9 to 32 Ave. 15.5	4 to 18 Ave. 8.8	85.7	91.9	Tower 1 and 2 at depth 32 feet with intermediate clarifier
9/1/77 to 9/9/77	5	25 to 26 Ave. 25.4	66 to 100 Ave. 85.6	4 to 17 Ave. 13.0	5 to 14 Ave. 10.0	84.8	88.3	Tower 1 and 2 at depth 32 feet without intermediate clarifier
9/12/77 to 11/23/77	17	14 to 25 Ave. 19.3	62 to 384 Ave. 138.1	9 to 44 Ave. 20	6 to 42 Ave. 18.9	85.5	86.3	Tower 1 at depth 16 feet, Tower 2 at depth 32 feet without intermediate clarifier.
3/29/78 to 4/28/78	20	12 to 15 Ave. 13.3	40 to 249 Ave. 139.5	4 to 47 Ave. 16.9	10 to 24 Ave. 16.2	87.9	88.4	Tower 1 at depth 16 feet, Tower 2 at depth 32 feet without intermediate clarifier.

tions ranging from 80 to 600 mg/1 (4). So the BOD removals were more than 85% in the Hoboken pilot plant study by using the rotoscreen as a primary treatment.

Due to river inflow and infiltration, a high chloride concentration, 50 to 770 mg/1 is found in the influent to the Hoboken Treatment Plant. However, no deleterious effect on the process was observed during the pilot test period.

The effect of NaCl on biological film has been reported by Lawton (5). He found that a step increase in salt concentration to 20,000 mg/1 had a deleterious effects on the film growth, but the film growth recovered in one day.

As shown in Table 1, the effluent BOD concentration from Tower 2 was lower than from Tower 1, but with two towers, the trickling filter did not significantly further reduce the BOD concentration. This was because some of influent BOD concentration were lower and the effluent BOD from Tower 1 had reached the equilibrium BOD. Therefore, the BOD removal efficiency was improved a little by increasing the number of towers, under this, wastewater characteristics and hydraulic loading rate. The provision of the additional tower to the treatment can secure the quality of the effluent BOD concentration. In case the influent BOD concentration is higher, the effluent of the first tower can be treated by the second tower.

The overall suspended solids removals in this study ranged from 67 to 82% as shown in Table 2. The effluent suspended solids concentration from two-stage trickling filter was less than 20 mg/1 after 30 minute settling.

Table 2

Effects of Wastewater Temperature, Intermediate Clarifier and Tower Depth on SS Removal of Plastic Trickling Filters

Period	Number of Samples	Effluent SS (mg/l)			SS Removal (%)		Note
		Screen	Tower 1	Tower 2	Tower 1	Overall	
8/15/77 to 8/31/77	12	22 to 72 Ave. 43.7	2 to 22 Ave. 13.7	2 to 17 Ave. 7.9	69	82	Tower 1 and 2 at depth 32 feet with intermediate clarifier.
9/1/77 to 9/9/77	4	37 to 75 Ave. 57.0	17 to 24 Ave. 18.6	9 to 22 Ave. 15.9	67	72	Tower 1 and 2 at depth 32 feet without intermediate clarifier.
9/12/77 to 11/23/77	12	35 to 100 Ave. 53.8	8 to 38 Ave. 23.3	5 to 51 Ave. 17.7	57	67	Tower 1 at depth 16 feet, Tower 2 at depth 32 feet without intermediate clarifier.
3/29/78 to 4/28/78	6	9 to 82 Ave. 58.2	8 to 37 Ave. 19.1	4 to 32 Ave. 18.9	67	68	Tower 1 at depth 16 feet, Tower 2 at depth 32 feet without intermediate clarifier

Rotating Biological Contactor

In this study, another configuration was made by using a rotating biological contactor (RBC) in place of the trickling filter to treat wastewater which was primarily screened by a rotoscreen. The treated effluent from the RBC, then, was settled by LAMELLA gravity settler. RBCs like trickling filters are a fixed-film biological treatment process. The settleable solids which passed through the rotoscreen would not effect the RBC. The reasons are explained in the section on trickling filters. The incoming settleable solids, which will not plug the media, may be removed in a final clarifier or LAMELLA gravity settler. The low BOD and SS effluent concentration shown in Tables 3, 4 and 5 implied the feasibility of flow-through system by using a rotoscreen for primary treatment and an RBC for secondary treatment. Therefore, using an RBC or a trickling filter in the flow-through system resulted the same efficiency on both BOD and SS removal.

The rotating bio-disc (or RBC) was arranged for two-stage operation at a hydraulic loading rate 2.5 gpd/ft^2 initially. The bio-disc was operated during two periods. One was from 8/12/77 to 12/8/77. The other was from 4/6/78 to 4/30/78. The result of the two periods are shown in Tables 3 and 4. The wastewater temperatures in the first period ranged from 12° to 26°C averaging 19.6°C. The major difference in the second period was the wastewater temperature ranged from 12° to 15°C averaging 14°C. The results from these two testing periods indicates that at a hydraulic loading of 2.5 gpd/ft^2 with a rotation rate of 1.6 rpm the two-stage bio-disc would obtain 84% BOD_5 removal and an expected BOD_5 effluent of 17.5 mg/l. Suspended solids levels in the effluent are expected to be 18 mg/l.

TABLE 3

BOD_5 and Suspended Solids Removal by Rotating Biological Contactor during 8/12/77 to 12/8/79

Parameter	Effluent Range	Average
BOD_5*	4 to 48 mg/l	18 mg/l
Suspended Solids*	1 to 44 mg/l	16 mg/l
BOD_5 Removal	61 to 100%	80%
SS Removal	-	60%

TABLE 4

BOD_5 and Suspended Solids Removal by Rotating Biological Contactors During 4/6/78 to 4/30/78

Parameter	Effluent Range	Average
BOD_5*	6 to 23 mg/l	15 mg/l
Suspended Solids*	1 to 57 mg/l	22 mg/l
BOD_5 Removal	75 to 95.7%	87.8%
SS Removal	-	66%

* After 60 minute laboratory settling for BOD analysis
After 30 minute laboratory settling for SS analysis

Table 5

Stage Analysis of RBC

Number of Stages	Period	HLR (gpd/ ft2.	Ave. Temper- ture (°C)	Number of Samples	Screen Effluent BOD (mg/1)	Screen Effluent TSS (mg/1)	Final Stage RBC Effluent MLSS (mg/1)	Final Stage RBC Effluent TSS (mg/1)	Final Stage RBC Effluent BOD (mg/1)	BOD Removal (%)	Net Solids Production	Sludge Production per BOD Removal (lb/lb)
2	4/6/78 to 4/30/78	2.5	14	19	40 to 249 Ave. 140.3	5 to 149 Ave. 65.5	4 to 136 Ave. 48.8	21.4	6 to 23 Ave. 15.4	89	-17	0.22
3	4/6/78 to 4/30/78	2.5	14	19	40 to 249 Ave. 140.3	5 to 149 Ave. 65.5	26 to 144 Ave. 59.1	21.6	9 to 25 Ave. 17.4	88	-6	0.31
4	7/19/78 to 7/28/78	2.5	24	8	59 to 180 Ave. 129	67 to 98 Ave. 81.7	31 to 171 Ave. 89	14	10 to 26 Ave. 21.3	83	7	0.70
5	7/19/78 to 7/28/78	2.0	24	8	59 to 180 Ave. 129	67 to 98 Ave. 81.7	41 to 184 Ave. 92.1	14	16 to 29 Ave. 21.3	83	9	0.73

In stage analysis, the RBCs were arranged for a two, three, four and five stage operation at hydraulic loading rates 2.0 and 2.5 gpd/ft^2. The RBCs were rotating at 1.7 rpm for the period of test. At two stage operation, the BOD removal was 89% as shown in Table 5. As the stages increased from 3 stages to 5 stages, the BOD removals were changed from 88% to 83%. The BOD removal efficiency versus stages is plotted in Figure 5. From the results, it was found the BOD removal was higher at two stages than any other additional stages. The performance of the RBC system was related to the characteristics of the wastewater, as it left the rotoscreen. The higher influent BOD in two and three stages might result in higher BOD removal efficiency. Although, on the average the five stages did not remove any further BOD, it did remove additional ammonia nitrogen. The four stage system on the average lowered the ammonia nitrogen concentration from 10.4 mg/1 to 5.2 mg/1, which is a 50% reduction. An additional 1.8 mg/1 ammonia nitrogen was removed by the fifth stage for a total ammonia nitrogen reduction in the five stages of 67.3%. Apparently, BOD reduction in terms of soluble BOD was nearly complete at a loading of 2.5 gpd/ft^2.

In terms of net solids production, the two stage system loaded at 2.5 gpd/ft^2 gave a net solids production of -17 mg/1; but to three and four stages gave -6 mg/1 and 7 mg/1 at 2.5 gpd/^{2}, to five stages gave 9 mg/1 at 2.0 gpd/ft^2 as shown in Table 5. The net solids production was calculated by substracting the suspended solids concentration in the rotoscreen effluent from the mixed-liquor suspended solids concentration in the RBC. This difference represents the suspended solids concentration changes through the RBC. The sludge production per BOD removals for various stages are shown in the last column of Table 5, and also plotted in Figure 6. Sludge production was calculated by substracting the suspended solids concentration in the final clarifier effluent from

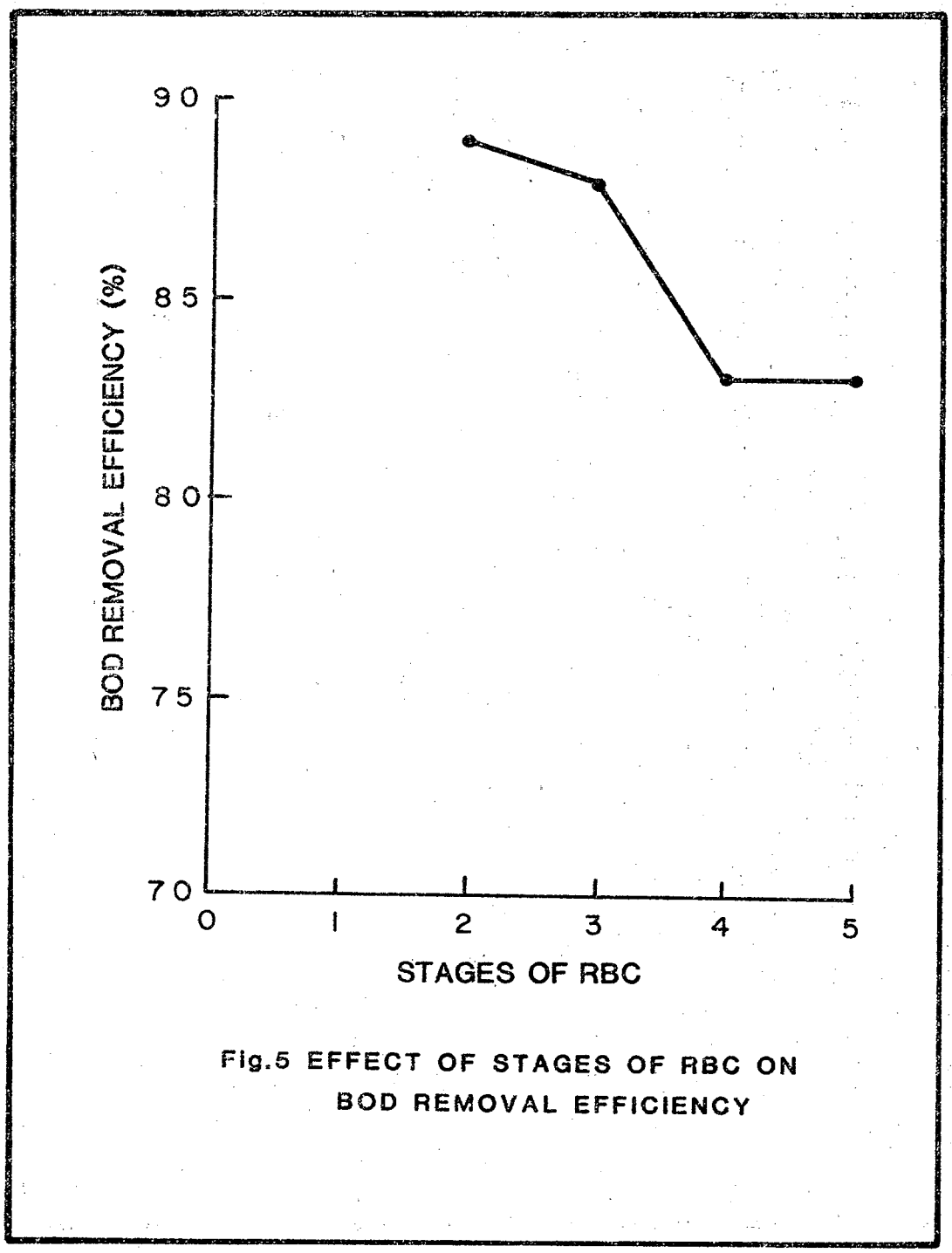

Fig.5 EFFECT OF STAGES OF RBC ON BOD REMOVAL EFFICIENCY

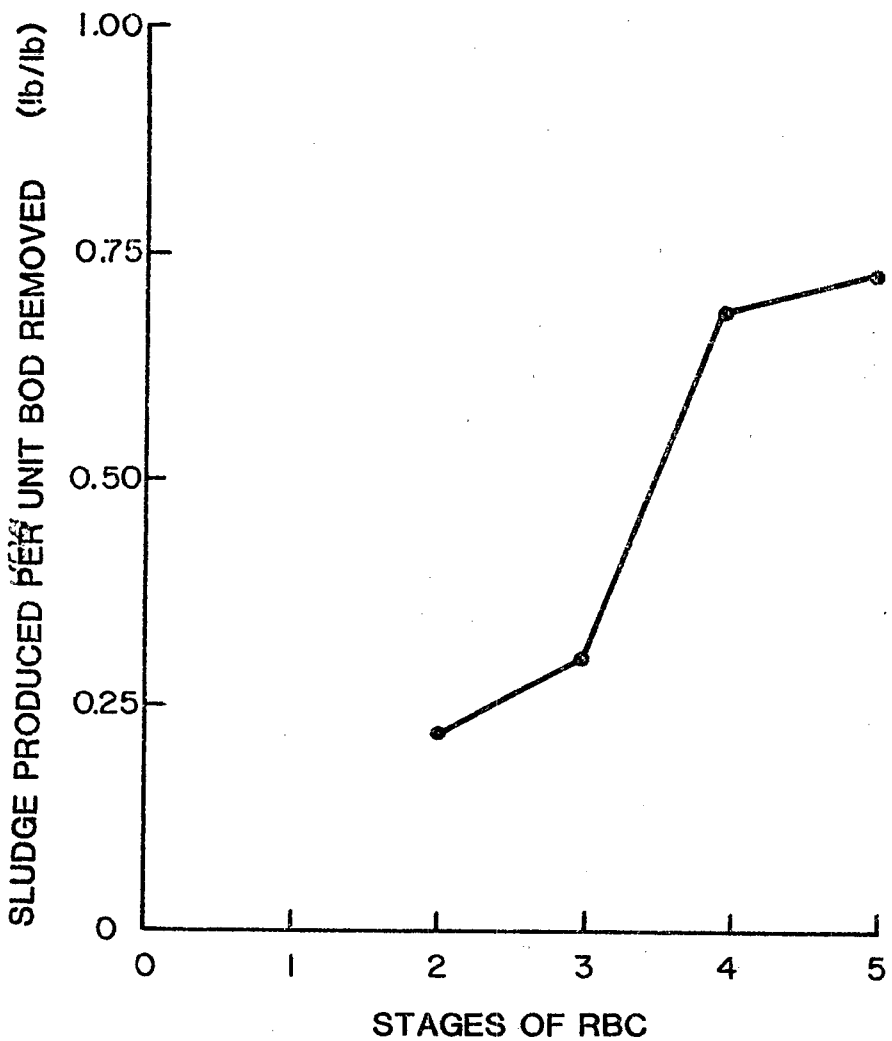

Fig.6 EFFECT OF STAGES OF RBC ON SLUDGE PRODUCTION PER UNIT BOD REMOVAL

the mixed liquor suspended solids concentration in RBC. Dividing this sludge production by the decrease in BOD concentration through the test unit yields sludge production as sludge produced per unit BOD removed. At the same hydraulic loading, the smaller number of stages produced low sludge production per BOD removal. This is exactly what would be expected from considerations of F/M where F is food and M is biomass. In the RBC system, the F/M ratio is directly to the F/A ratio, where A is the area of active surface area in the system. The more area in the first stage of the RBC system, the lower is its F/A ratio and by comparison with F/M, the lower should be the net sludge production. When the hydraulic loading rate increased from 2.5 gpd/ft^2 to 5.0 gpd/ft^2 in a two-stage RBC, the net sludge production per BOD removal would be more as expected F/M ratio increase. As shown in Table 6, at the two stages of RBC the net solids production and sludge production per BOD removal at 5.0 gpd/ft^2 are larger than that at 2.5 gpd/ft^2.

This observation of the performance of two RBCs versus a larger number in a staged RBC system gives one the feeling that future designs would be optimized by spreading the surface area over a smaller number of stages. For carbonaceous organic removal, this would lead to a lower sludge production, a more stable biomass, and a lower BOD in the final effluent caused not by better soluble BOD removal, but more by the lower BOD of the more stable biomass that might escape settling in the final clarifier. The optimum number of stages would appear to be two stages, for BOD removal and for a given amount of tital supplied surface area. When comparing with single state system, two stage would reduce the impace of short circuit, toxic, and shock loading due to low F/M.

LAMELLA Gravity Settler

The LAMELLA gravity settler was used for final clarification in the two treatment configurations of pilot plant study. The testing results are shown in Tables 7 and 8.

The hydraulic variation imposed on the LAMELLA gravity settlers included loadings of 360, 720 and 1008 gpd/ft^2 in the rotating bio-disc study. The results are shown in Table 7. During the 360 gpd/ft^2 loading priod the effluent quality averaged 14 mg/l; the corresponding removal averaged 80.5%. This removal efficiency reflects the excellent floc development in the RBC process. However, the 360 gpd/ft^2 loading provided no real challenge for the LAMELLA gravity settler. Correspondingly, the hydraulic loading rates were increased. At the 720 and 1008 gpd/ft^2 loading the effluent SS was 23 and 14 mg/l respectively. Even the effluent SS were lower than 30 mg/l, however, the removal efficiency were not shown as good. This was due to the lower solids flux through the rotating biodisc process during the study.

In a two-day investigation, it was found the effluent SS was 12 mg/l at the 1580 gpd/ft^2 loading. The results indicate the LAMELLA gravity settler can treat rotating bio-disc effluent at higher surface loading rate. The intent of this quick study was to examine dramatically higher loadings while the plant was relatively stable. Althouth the LAMELLA gravity settler displayed good suspended solids removals at the 1580 gpd/ft^2 load

ing, this test was too short to be utilized for full scale design. However, higher design loading 720 or 1008 gpd/ft^2 are probable.

The performance of LAMELLA gravity settler for the effluent of the trickling filter process which was operated in the 1.4 gpm/ft^2 is shown in Table 8. The average effluent suspended solids was 8 and 11 mg/1 while the loading rate on the LAMELLA gravity settler was 360 and 1008 gpd/ft^2 respetively. These results exhibited that even when the loading rate was up to 1008 gpd/ft^2 the effluent suspended solids was 11 mg/1 and removal effiency was 75%.

Therefore, the investigation indicates that the LAMELLA gravity settler is suitable for final solids separation following either a rotating bio-disc or trickling filter system. The loading rates on the LAMELLA gravity settler can be up to 1008 gpd/ft^2. The investigation also found the effect of biological solids on the intervals of the LAMELLA gravity settler was minor. Some growth occurred on the wetted surfaces which only appeard as a film and never affected the workings of the LAMELLA gravity settler. The LAMELLA gravity settler was cleaned twice in the six months of pilot operation and only then because of changeovers in operations. Clogging problems are existant in waste treatment application(2). But the application of LAMELLA gravity settler on rotating bio-disc and trickling filter solids exhibited a good solids separation. The effluent solids of the rotating bio-disc and trickling filter are unlike those produced by activated sludge. The good settling characteristics of the solids found in the rotating bio-disc or trickling filter effluent is due to the low mixed liquor suspended solids concentration and result in discrete settling which greatly enhance the solids settling velocity. Zone, hindered or compression,

Table 6

The Effect of Hydraulic Loading Rate on Net Solids Production and BOD Removal at Two-Stage RBC

Period	Temperature (°C)	HLR (gpd/ft^2)	Net Solids Production (mg/1)	BOD Removal (%)	Sludge Production per BOD Removal (lb/lb)
7/19/78 o /28/78	24	5.0	21	82.2	0.81
4/6/78 to 4/30/78	24	2.5	1.7	85.0	0.50

Table 7

The Performance of Lamella Gravity Settler for the Effluent of Bio-Disc Process

Overflow Rate (gpd/ft^2)	Parameter	Effluent (mg/1)	Overall Removal (%)
360	BOD_5	14	80.5
	SS	9	82.4
720	BOD_5	28	75.0
	SS	23	60.0
1008	BOD_5	21	84.5
	SS	14	67.5

settling does not occur in the secondary clarifier following the biological contactor system. The removal of discrete particles is independent of tank depth and detention time and is a function only of the overflow rate. In plate settler the flow is laminar and uniformly distributed. The suspended solids are not upset by outside forces such as convection currents or sudden hydraulic change.

Microstrainer

A 35 micron microstrainer was directly applied to the effluent from the two-stage RBC with hydraulic loading rate of 2.5 gpm/ft^2. The test results of BOD and SS removal of two-stage RBC system, which received wastewater from a roto screen, by microstrainer and simulated clarifier are shown in Table 9. The simulated clarifier is mentioned before when settling RBC effluent in a one-liter graduated cylinder for 60 and 30 minutes for BOD and SS tests respectively.

The results in Table 9 indicate that a microstrainer achieved better SS removal than the simulated clarifier, while the effluent BOD_5 was ony 6 mg/l. However, after August 31, 1977 it was found that the microscreen failed due to heavy slime growths. Microorgamisms grew up quickly on the screen's surface which caused clogging and biofloc in the effluent. The use of chlorinated water for cleaning gave only temporary improvement. Those operation difficulties made it impossible to apply a microstrainer for the removal of SS from RBC effluent.

Table 8

The Performance of Lamella Gravity Settler for the Effluent of the Trickling Filter Process

Overflow Rate (gpd/ft^2)	Parameter	Effluent (mg/l)	Overall Removal (%)
360	BOD_5	12	90
	SS	8	81
1008	BOD_5	22	86
	SS	11	75

Table 9

BOD and SS Removals of Two-Stage RBC System by Microstriner and Simulated Clarifier

Parameter	No of Samples	Ave. Effluent Conc. at RBC (mg/l)	Ave. Effluent Conc. at Microstrainer (mg/l)	% removal by Microstrainer	Ave. Effluent at simulated Clarifier (mg/l)	% removal by simulated Clarifier
BOD	12	106	6	94.3	11	80.6
SS	12	44	8	81.9	10	77.3

Test Period: From 8/15/77 to 8/31/77

North Bergen Treatment Plant

The concept of design for the North Bergen Central Sewage Treatment Plant was based on the results of Hoboken pilot plant study which was proven to be a reliable technology. This 10 MGD treatment plant was finished in 1981. According to the operation report of North Bergen Treatment Plant, the average data for the first 4 months, October, November, and December of 1981 and January of 1982 are summarized in Tables 10, 11 and 12. The numbers shown on Tables 10, 11 and 12 are the averages of data, based on daily sampling. Table 10 shows the flow rates and characteristics of raw wastewater. The actual average flow was from 0.85 to 1.5 MGD which is much less than the design flow rate 10 MGD. This is because all areas have not been hooked up the sewer line yet. Therefore, part of the facilities at North Bergen Central Sludge Treatment Plant are not in operation.

The application of rotoscreens, in the North Bergen Central Sewage Treatment Plant, was for primary treatment. The results of the operation are shown in Table 11. The suspended solids removals were from 24.9 to 33.6% averaging 28.9% during the first four month operation. The manufacturer's information showed about 15 to 20% SS removal when applying rotoscreening to municipal sewage. Therefore, 20% SS removal by applying rotoscreens for preliminary treatment can be expected. Two advantages of rotoscreens are high dry weight of solids produced and small floor requirements, associated with low operation and capital costs. The maintenance costs are also lower due to fewer moving parts when compared with primary clarifiers.

Two-stage rotating biological contactors were applied to remove BOD in the North Bergen Central Sewage Treatment Plant. The BOD removal efficiency ranged from 84.1 to 89.7% averaging 86.8% during the first four months. Compared with the data in

Table 10

Raw Wastewater Flow and Characteristics
of North Bergen Central Sewage Treatment Plant

Month 1981 and 1982	Ave. Flow (MGD)	Temperature (°C)	pH	BOD (mg/1)	SS (mg/1)	NH_3-N (mg/1)
October	1.5	21.7	7.4	125	132	-
November	0.942	20.2	7.48	130.5	97.7	14.6
December	0.849	16.1	7.29	95.5	97	18.8
January	0.981	14.6	7.46	85.5	102.3	15.5

Table 11

Average SS Removal by Rotoscreen in the North Bergen Central Sewage Treatment Plant

Month (1981 and 1982)	Influent SS (mg/1)	Effluent SS (mg/1)	SS Removal (%)
October	133.3	92.3	30.8
November	109.3	72.6	33.6
December	134.2	100.8	24.9
January	96.1	71	26.1
Average	118.2	84.2	28.9

Table 12

Average BOD, SS and NH_3-N Removals of
North Bergen Central Treatment Plant

Month (1981 and 1982)	Ave. BOD, mg/l			Ave. SS, mg/l			Ave. NH_3-N,		
	Inf.	Eff.	% Removal	Inf.	Eff.	% Removal	Inf.	Eff.	% Removal
October	125	12.9	89.7	132	22.5	83.0	-	-	-
November	130.5	20.7	84.1	97.7	22.4	77.1	14.6	8.3	43.2
December	95.5	14.3	85.0	97	18.5	80.9	18.8	6.97	62.9
January	85.5	9.9	88.4	102.2	11.4	88.9	15.5	5.4	65.2
Ave.	109.1	14.5	86.8	107.3	18.7	82.5	16.3	6.9	57.1

Hoboken pilot plant study, the average influent BOD concentration was 109.1 mg/l which is about the same influent BOD concentration at the Hoboken Pilot Plant study. The average effluent BOD concentration was 14.5 mg/1 which is close to or a little lower than that in the Hoboken pilot plant study. However, nitrification was already ocurring in the North Bergen treatment plant. In Table 12, it shows the averaging ammonia nitrogen removal was 57.1%. The average ammonia nitrogen concentration was reduced from 16.3 mg/l to 6.9 mg/l.

Suspended solids removal ranged from 77.1 to 88.9% averaging 82.5%. The average effluent, suspended solids concentration of 18.7 mg/l from LAMELLA gravity clarifier implied that the settlability of biological slime sloughed from RBC was good. After four months operation the biological solids did not grow on the surfaces of LAMELLA gravity plate which also implied that the characteristics of biological solids from RBC was suitable to the use of LAMELLA gravity settler for the final liquid solids separation.

Some advantages of LAMELLA gravity settlers includes low space requirements, low installed costs, low maintenance, due to fewer moving parts to wear, replace and adjust, and high efficiency. In a well designed and properly sized LAMELLA gravity settler, the flow is laminar, therefore, the suspended solids are not upset by outside forces such as convention currents or sudden hydraulic change.

The successful operations in the North Bergen Central Sewage Treatment Plant during the last four months implied that the application of flow-through system to municipal sewage is feasible. The short

detention time of rotoscreening in the system provides the great advantage to the biological treatment system. Usually, the hydraulic detention in the primary clarifier is two hours. But the detention in the rotoscreen is only 2 to 3 minutes. Therefore, the difference between the influent temperature and the effluent temperature is not too much. This is very important for biological activity during the winter time. The higher water temperature makes the biological activity in the reactor higher. The comparison of rotoscreen versus primary settling tank and LAMELLA gravity settler versus secondary settling tank to their detention times and land requirements are listed in Table 13. The detention time in the entire system including rotoscreen, RBC, and LAMELLA gravity settler takes less than an hour, compared with the detention time of 6 hours in conventional activated sludge. This compact system also provides another benefit, due to its compactness, the treatment system in the North Bergen treatment plant is housed in a building which protects the rotoscreen, RBC, and LAMELLA gravity settler equipment from extreme temperature, variation, heavy rain and high wind. The construction cost of RBC using shallow tanks (6 ft. above ground) versus activated sludge tank with 15 foot depth underground is relatively lower as the water in the activated sludge tank exerts considerable pressure on the soil, thereby requiring costly pile foundation and dewatering, especially, if the soil conditions on site are bad. The total construction cost of the North Bergen Central Sewage Treatment Plant in 1980 was 9.5 million dollars. An equivalent secondary treatment plant would have cost at least 22 million dollars, (figures obtained by using data on the Constructions Costs for Municipal Wastewater Treatment Plants (1978) (6) updated to 1980 dollars). The low operation and maintenance costs, minimum land requirements and reduced capital cost inherent in this flow-through system show great potential in the treatment of municipal sewage, in the near future.

Table 13

Comparison of Detention Times and Land Requirements
Between Conventional Activated Sludge and
North Bergen Treatment Plant

Conventional Activated Sludge			North Bergen Treatment Plant		
Units	D.T. (Min.)	Land (ft^2)	Units	D.T. (Min.)	Land(ft^2)
Primary Settling Tank	90 to 150	17,000	Rotoscreen	2 to 3	480
Conventional Activated Sludge	360	26,000	Rotating RBC	30 to 60	12,320
Secondary Clarifier	90 to 150	17,000	Lamella Gravity Settler	5 to 10	1,600

Design Flow = 10 MGD

SUMMARY AND CONCLUSIONS

The Hoboken pilot plant was operated on a full-scale facility by using rotoscreens for primary treatment. Rotating biological contactors or trickling filter for carbonaceous removal, and LAMELLA gravity settler for biological solid and liquid separation. The results showed this flowthrough system can successfully treat sewage at the effluent concentration of both BOD and suspended solids less than 20 mg/l. The biomass sloughed from the media of rotating biological contactor or trickling filter into the mixed liquor were easily settled down on the LAMELLA gravity settler without causing any problem on the settler, such as biological growth on the plater of the settler.

In stage analysis, the operation of the rotating biological contactor at two stages showd the best results, either in the BOD removal efficiency or sludge production, while comparing with 3, 4 or 5 stages.

Therefore, the concept of flow-through system and two-stage rotating biological contactor was adopted in the design of the North Bergen Central Sewage Treatment Plant.

From the operation data at the first four months, it showed the flow-through system was successfully applied to the North Bergen Central sewage Treatment Plant. The rotoscreen removed the suspended solids from 24.9 to 33.6% averaging 28.9% which is less than that of the primary settling tank. However, the effluent of the rotoscreen would not effect the whole process. The average BOD and SS concentration form LAMELLA gravity settler runs 14.5 mg/l and 18.9 mg/l respectively. Due to the short hydraulic detentioin time of 2 to 3 minutes at the rotoscreen the temperature between influent and effluent would not make much difference. This short detention time provides a

great advantage in biological treatment, in which the biological activity is temperature relative, during the winter weather. The low construction cost of 11 million dollars versus 22 million dollars for the equivalent secondary treatment plant, low operation and maintenance cost, low land requirements and low energy cost implied the flow-through system a great potential to the treatment of municipal sewage in the near future. When the physical conditions are limited by available land size, poor soil conditions, and high groundwater levels, the authors recommend that you should consider the application of this flow-through system for municipal wastewater treatment.

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AN IN DEPTH COMPLIANCE AND PERFORMANCE ANALYSIS OF THE RBC PROCESS AT MUNICIPAL SEWAGE TREATMENT PLANTS IN THE UNITED STATES

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INTRODUCTION

A number of recent governmental reports, most notably the Government Accounting Office (G.A.O.) report entitled, "Costly Wastewater Treatment Plants Fail To Perform As Expected", have severely criticized the performance efficiency of existing wastewater treatment plants. The plants evaluated by the G.A.O. were reported to employ either activated sludge, trickling filter or lagoon unit operations for biological wastewater treatment. Autotrol Corporation, as part of a routine customer service, regularly checks the operational performance efficiency of Rotating Biological Contactor (RBC) plants where it has supplied equipment. This paper statistically compares the performance efficiency of those treatment plants evaluated by the G.A.O. with RBC plants supplied by Autotrol Corporation. This report was originally written early in 1981 and further modified after evaluating an additional 22 plants early in 1982. It should be noted that the data from 1980 and 1981 was statistically similar and 1981 data verified original conclusions presented in the initial report.

In general, while the G.A.O. report stated that "E.P.A statistical report on plant performance show that between 50 and 75%

of the treatment plants in operation are violating their permits in any given time", Autotrol's study on performance of existing RBC plants revealed that these RBC plants were meeting their discharge requirements for BOD removal 89% of the time and were meeting their discharge requirements for suspended solids 92.5% of the time. The data further suggests that long term process failures (violations occurring more than six months per year) is 500% more likely in alternative technology evaluated by G.A.O. as compared to RBC plants evaluated by Autotrol.

When discharge violations did exist in RBC facilities, the vast majority of the violations were minor excursions from standards. Where major violations occurred, the prime cause was equipment deficiency, not treatment reliability.

DESCRIPTION OF THE STUDY

In November of 1980, the General Accounting Office reported back to the Congress the results of its investigation into the operation of Publicly Owned Treatment Works (POTW) that were designed, built, and funded under the authority of the grants program. The report stated that at any given time 50% to 75% of the plants are in violation of their National Pollutant Discharge Elimination System (NPDES) permits. The report continues that a random sample of 242 plants in 10 states found 87% of the plants in violation of their permits at least one month per year with 27% in "serious" violation.

Autotrol Corporation decided to statistically analyze and compare the results of its RBC facilities to those analyzed by the G.A.O. Important to the confidence of the reader is the knowledge that all the data displayed and the only data we would use is that gathered by the municipality and reported to its appropriate state agency.

Although every attempt was made to make the comparison as similar as possible, i.e.,

> Autotrol and G.A.O. evaluated plants classified by E.P.A. as capable of providing secondary or better levels of treatment.
>
> Autotrol and the G.A.O. evaluated plants based on the issued NPDES permit.

Some differences that did exist are as follows:

G.A.O. evaluated 242 plants from the universe of 676 plants available for evaluation (36%). Autotrol evaluated 46 plants out of a possible 160 plants (29%).

The G.A.O. plants were "randomly selected". Autotrol plants are those for which sufficient data was available to make a meaningful analysis and comparison.

The G.A.O. report selected plants of flow ranges between 1.0 mgd to 50.0 mgd. Autotrol plants had a flow range of between 0.2 to 8.0 mgd.

G.A.O. evaluated data for a one year period between 1978-1979. Autotrol evaluated annual data for 1980 and for 1981.

The G.A.O. report evaluated data for BOD_5, TSS and fecal coliform. Because of the limited amount of data on fecal coliform, Autotrol's evaluations were based on BOD_5 and TSS only.

RESULTS

The summary of plants evaluated, the violations observed, and the number of monthly sample periods are described in Table I for the 1980 data base and in Table II for the 1981 data base. The following charts show the compilation of both 1980 and 1981 data for all monthly data reviewed and the resultant successful compliance ratio.

PERFORMANCE CHARACTERISTICS OF RBC PLANTS
SAMPLE PERIOD 1980-1981

	Number of Monthly Samples	Number of Monthly Violations	% Successful Operation
	BOD_5		
All Plants Surveyed	598	66	88.96%
	TSS		
All Plants Surveyed	598	45	92.47%

The results indicate that regardless of load and flow conditions, plant operation, or potential lab analysis error, the RBC plants evaluated were meeting discharge requirements 89% of the time for BOD removal and 92.5% of the time for TSS removal. This high degree of compliance is considered very good, particularly in light of the G.A.O. report statement, "E.P.A.'s statistical reports on plant performance show that between 50 and 75 percent of the treatment plants in operation are violating their permit at any given time".

In an attempt to directly correlate the G.A.O. report on the number of plants with discharge violations with the Autotrol report on RBC facilities, Tables III, IV and V were developed. Table III reports the number of plants evaluated in various regions by the G.A.O., the total number of plants in violation and the number of these violations per year of operation. Autotrol developed similar comparisons for the 1980 sample period (Table IV) and for the 1981 sample period (Table V). In addition, because certain plants (#3, 18, 33, 44, 45) had data for both 1980 and 1981, worst year data for duplicate plants were used to develop a composite summary of RBC facilities. The composite RBC plant performance is shown in Table VI.

The following conclusions can be drawn from a comparison of Tables III, IV, V and VI:

1. The plants evaluated by G.A.O. demonstrated that 49.2% of the plants violated discharge permits for more than 6 months out of the year. Less than 11% of Autotrol RBC plants had these extended violations of more than 6 months per year. (Table III versus Table VI).

2. While 25.6% of the plants evaluated by G.A.O failed to achieve successful operation for more than 3 months per year, only 4% of the RBC plants surveyed performed this poorly. (Table III versus VI).

3. While in excess of 87% of the G.A.O. plants experienced at least minor violations, only 37% of the RBC plants experienced similar difficulties. That is, 63% of all RBC plants continually met discharge limits month in and month out. (Table III versus VI).

4. While only 34.7% of the G.A.O. plants surveyed performed satisfactorily for more than 9 months per year, the RBC data showed that 82.6% of the plants performed satisfactorily for periods in excess of 9 months per year.

5. The data developed for RBC plants were conducted over a 2-year period. The comparison between 1980 and 1981 data basis was statistically similar and repetitive.

6. The chances of having no violation in any year is 4 times greater using RBC plants evaluated by Autotrol as compared to employing those plants evaluated by G.A.O.

7. The chances of long term process failure (i.e. greater than 6 months) is 5 times as likely in those plants evaluated by G.A.O. as compared to Autotrol RBC facilities.

DISCUSSION OF VIOLATIONS OF RBC FACILITIES (COMPOSITE 1980-1981 DATA

Of the forty-six (46) RBC plants evaluated, seventeen (17) plants experienced violation of wastewater NPDES permits. Twelve (12) can be classified as non-serious violations while five (5) can be termed serious violations. The definition of serious violation is as defined in the G.A.O. report, "when one or more of the three parameters was violated for more than four (4) consecutive months during the review period and averaged more than 50% above the permit limit during the period of non-compliance". The definition of serious violation is similar in Autotrol's evaluation with the exception being that fecal coliform data was insufficient as a parameter to be evaluated. The following discussions will classify the extent of non-serious and serious violations encountered during our survey.

NON-SERIOUS VIOLATIONS

Of the twelve (12) plants experiencing non-serious violations, nine (9) of those facilities had yearly average discharge BOD_5 and TSS values lower than their monthly permit allowance. (In general, the yearly average BOD_5 discharge value was 75% of the monthly discharge requirement). Of those plants that exceeded the discharge requirement on a yearly average basis, a highest yearly

average value of 37 ppm BOD_5 and a highest yearly average TSS value of 32 ppm was recorded. In general, a review of those plants experiencing non-serious violations indicate that violations are minor, of very short term, and appear to have been corrected in the most recent operations.

SERIOUS VIOLATIONS

Five (5) plants in our survey experienced serious violation (10.9%). This compares to the G.A.O. survey where sixty-six (66) plants (27.3%) had serious violations. Two (2) of the five (5) RBC plants experienced mechanical problems which resulted in poor performance. One of the five facilities experienced industrial waste loads with inadequate pretreatment which resulted in inferior performance. It is not known whether the mechanical problems were caused by design or equipment deficiencies. The following chart describes the plant, reason for non-compliance, average effluent BOD_5 values and average effluent yearly TSS yearly values for the survey year.

Plant	Major Category	Avg.Yearly Eff. BOD_5	Avg.Yearly Eff. TSS
#11 WI	Industrial Waste Overload	61 ppm	40 ppm
#24 OH	Equipment Deficiency	15 ppm	3 ppm
#41 IA	Equipment Deficiency	41 ppm	42 ppm
#34 WA	O&M Deficiency	42 ppm	19 ppm
#22 WI	O&M Deficiency	8 ppm	26 ppm

It should be noted that even though these plants experienced serious violations, two of the five facilities still provided effluent quality classified as better then secondary treatment by the E.P.A.

CONCLUSIONS

The above comparative analysis indicates that RBC facilities performed significantly better than facilities evaluated by the G.A.O. Conformance was better in terms of both non-serious and serious violation categories.

TABLE I

MONTHLY NPDES VIOLATIONS DURING 1980

Plant No.	State	$TBOD_5$ Mo. Violation Per No. of Mo. Surveyed	TSS Mo. Violation Per No. of Mo. Surveyed
3	PA	1/12	0/12
4	OH	0/12	0/12
5	IN	0/12	0/12
6	MI	0/12	0/12
8	IA	0/12	0/12
10	MN	1/12	0/12
11	WI	12/12	12/12
13	WI	0/11	0/11
16	KY	0/12	0/12
17	MI	0/12	0/12
18	CO	0/12	1/12
19	OR	0/12	0/12
21	MI	0/12	0/12
23	CO	0/12	1/12
24	OH	7/12	0/12
26	MI	0/12	0/12
27	WI	0/12	0/12
28	NY	0/12	0/12
30	WI	0/12	0/12
31	WI	1/12	0/12
32	KY	0/11	0/11
33	WI	3/12	0/12
38	WI	2/12	2/12
39	KS	0/12	0/12
41	IA	8/12	9/12
44	NE	1/12	2/12
45	OR	1/12	0/12
46	WA	0/10	0/10
48	WA	0/11	0/11
TOTALS:		37/343	27/343
Compliance Ratio		89.2%	92.1%

TABLE II

MONTHLY NPDES VIOLATIONS DURING 1981

Plant No.	State	$TBOD_5$ Mo. Violation Per No. of Mo. Surveyed	TSS Mo. Violation Per No. of Mo. Surveyed
3	PA	1/11	0/11
5	IN	0/12	0/12
9	WI	6/11	0/11
12	WA	0/12	0/12
14	IL	0/12	3/12
16	KY	0/12	0/12
17	MI	0/12	0/12
18	CO	1/12	0/12
21	MI	0/12	0/12
22	WI	0/12	7/12
25	MI	0/11	0/11
26	MI	0/12	0/12
33	WI	2/12	1/12
34	WA	11/12	1/12
35	SD	0/12	0/12
36	IL	0/12	0/12
40	MI	0/12	0/12
44	NE	4/10	3/10
45	OR	0/12	1/12
46	WA	0/12	0/12
47	OR	0/10	0/10
50	WA	4/10	2/10
TOTALS:		29/255	18/255
Compliance Ratio		88.7%	92.9%

TABLE III

G.A.O. REPORT PLANT MONTHLY VIOLATIONS

G.A.O. Report Summary Region	Sample Number	Facilities in Violation				
		At Least 1 month	Number of Months			
			1-3	4-6	7-9	10-12
Boston	100	94	13	20	28	33
Chicago	92	74	23	15	13	23
San Francisco	50	43	17	4	16	6
TOTAL	242	211	53	39	57	62
%	100	87	21.9	16.1	23.6	25.6
Cumulative %			21.9	43.6	61.6	87.2
Inverse Cumulative %			87.2	65.3 (34.7)	49.2	25.6

TABLE IV

1980 SURVEY OF AUTOTROL RBC PLANT MONTHLY VIOLATIONS

Autotrol Report Summary Region	Sample Number	Facilities in Violation				
		At Least 1 Month	Number of Months			
			1-3	4-6	7-9	10-12
Various Across U.S.	29	12	9	0	2	1
%		41.4	31.0	0	6.9	3.5
Cumulative %			31.0	31.0	37.9	41.4
Inverse Cumulative %			41.4	10.5	10.5	3.5

TABLE V

1981 SURVEY OF AUTOTROL RBC PLANT MONTHLY VIOLATIONS

Autotrol Report Summary Region	Sample Number	Facilities in Violation				
		At Least 1 Month	Number of Months			
			1-3	4-6	7-9	10-12
Various Across U.S.	22	10	5	3	1	1
%		45.5	22.7	13.6	4.6	4.6
Cumulative %			22.7	36.4	40.9	45.5
Inverse Cumulative %			45.5	22.7	9.1	4.6

TABLE VI

COMPOSITE OF RBC PLANTS IN SURVEY - (1980 & 1981)

Autotrol Report Summary Region	Sample Number	Facilities in Violation				
		At Least 1 Month	Number of Months			
			1-3	4-6	7-9	10-12
Various Across U.S.	46	17	9	3	3	2
%		37.0	19.6	6.5	6.5	4.3
Cumulative %			19.6	26.1	32.7	37.0
Inverse Cumulative %			37.0	17.4 (82.6)	10.8	4.3

THE USE OF PLASTIC MEDIA TRICKLING FILTERS
TWO CASE HISTORIES

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INTRODUCTION

Over the last ten years, Jones & Henry Engineers has investigated the use of fixed film biological processes at a number of locations in the Great Lakes area. Approximately half of these studies have been directed toward the treatment of high strength waste and half toward the production of nitrified effluents. In some cases, the investigations have included rotating biological contactors and trickling filters.

This paper discusses two case histories on the use of plastic media trickling filters. One of the case histories deals with the treatment of high strength wastes at Kalamazoo, Michigan and the other with nitrification of the Lima, Ohio secondary effluent.

THE KALAMAZOO CASE HISTORY

The City of Kalamazoo, Michigan conditions sludge by wet air oxidation.[1] The conditioned sludge is thickened in decant tanks, dewatered by vacuum filters, and incinerated. The residue ash is landfilled. The decant tank supernatant and the vacuum filter filtrate are recycled to the head of the plant. Although the recycle streams represent less than 1 percent of the plant flow, they constitute 24 percent of the total organic load.

An alternative to direct recycling is treating the supernatant separately. Jones & Henry Engineers tested several processes for possible separate treatment of the wet air oxidation recycle streams as part of the design for advanced wastewater treatment. The practicality of various separate treatment processes was ascertained through desk top, bench scale, and pilot studies. Process effectiveness was judged on BOD reductions. Process viability was determined using additional paramaters including color removal, odor control, process reliability, operational simplicity, space requirements, and economics. Bench scale studies of activated sludge and physical-chemical methods showed these processes to be ineffective. Pilot plant investigations demonstrated that attached growth reactors would best be used for pretreatment of the recycle streams. Design information was developed during the tests that substantiated and expanded previous research.

CHARACTERISTICS OF RECYCLE STREAMS

The supernatant and filtrate have essentially the same physical and chemical properties. The coffee-colored liquors have an average temperature of 120°F (49°C), a noticeable odor, and strong frothing tendencies.

Standard chemical analysis of wet air oxidation by-product samples showed the liquor is acidic and extremely rich in nitrogen, with substantial amounts of chloride and sulfate and low levels of phosphorus and suspended solids. The waste has high BOD, COD, and TOC. Specific chemical characteristics are detailed in Table I.

THE STUDY

A plastic media trickling filter was pilot tested on the supernatant of the wet air oxidation decant tank. The unit was operated for nearly six months (December 17, 1974 to June 2, 1975). The temperature and oxygen demand of the supernatant were controlled by dilution with ground water. This was necessary as the temperature of the recycle stream was too high for biological treatment.

The test parameters for the pilot program were total and soluble BOD and COD, and total and volatile suspended solids. Color and odor were noted but not measured. The parameters were monitored on the influent and effluent to determine process efficiency at various organic and hydraulic loadings, and dilution ratios. Samples were collected seven days per week; one sample every two hours. Analyses were performed on daily composites of grab samples.

Figure 1 is a diagram of the pilot facility. The supernatant was diluted with ground water at the top of the tower. This mixture flowed through a funnel to a rotary distributor that controlled the hydraulic loading to the filter. The effluent was collected in the recycling reservoir at the bottom of the filter, and a portion of it was returned to a tank at the top of the tower. The constant level feed and recycle tanks were equipped with outlets to ensure constant discharge. The feed, recycle rates, and dilution ratio were varied to study the filter's efficiency under different conditions.

TABLE I

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

CHARACTERISTICS OF WET AIR OXIDATION SUPERNATANT

Test Parameter		Test Parameter	
HYDROGEN ION CONCENTRATION (pH @ 25°C)	4.65	CHEMICAL OXYGEN DEMAND (mg/l)	18,200
COLOR, TRUE (ALPHA Units @ 25°C)	7,600	TOTAL CARBON (mg/l)	7,800
TURBIDITY, APPARENT (JTU @ 25°C)	70	TOTAL ORGANIC CARBON (mg/l C)	7,785
ACIDITY, TOTAL (mg/l $CaCO_3$)	1,580	ALUMINUM, TOTAL (mg/l Al)	10.4
TOTAL SOLIDS (mg/l @ 103°C)	14,657	CALCIUM, TOTAL (mg/l Ca)	196
SUSPENDED SOLIDS (mg/l @ 103°C)	120	POTASSIUM, TOTAL (mg/l K)	248
SETTLEABLE SOLIDS (ml/l @ 60 minutes)	0.4	MAGNESIUM, TOTAL (mg/l Mn)	66.0
TOTAL KJELDAHL NITROGEN (mg/l N)	713	SODIUM, TOTAL (mg/l Na)	180
AMMONIA NITROGEN (mg/l N)	455	IRON, TOTAL (mg/l Fe)	6.28
NITRATE NITROGEN (mg/l N)	14.3	COPPER, TOTAL (mg/l Cu)	0.21

TABLE I
(Continued)

Test Parameter		Test Parameter	
PHOSPHORUS, TOTAL (mg/l P)	1.03	CHROMIUM, TOTAL (mg/l Cr)	1.74
CHLORIDE, TOTAL (mg/l Cl)	512	CADMIUM, TOTAL (mg/l Cd)	0.08
SULFATE, TOTAL (mg/l SO_4	445	LEAD, TOTAL (mg/l Pb)	0.22
BIOCHEMICAL OXYGEN DEMAND (mg/l BOD @ 5-days)	8,200	NICKEL, TOTAL (mg/l Ni)	0.76
		ZINC, TOTAL (mg/l Zn)	10.4

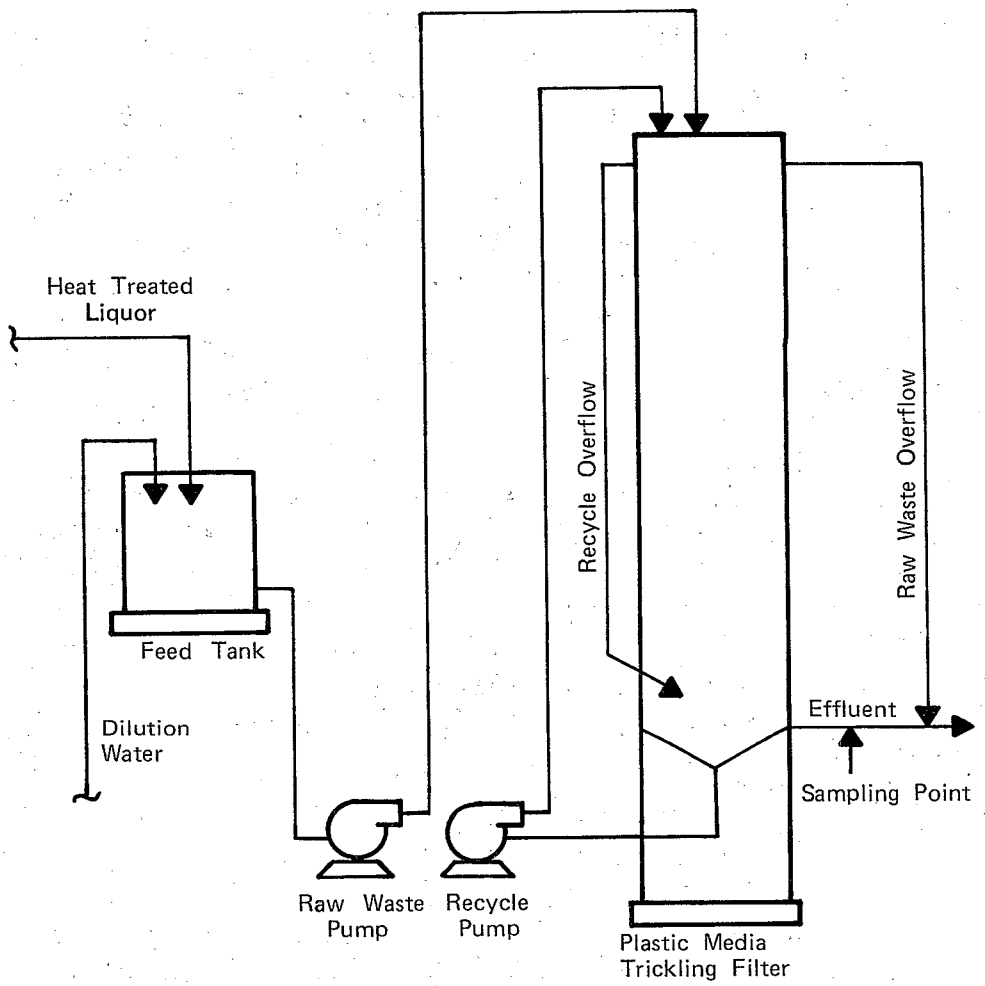

FIGURE 1

DIAGRAM OF PILOT FACILITY

The pilot equipment included:

Pilot Tower Steel shell	- 3 feet in diameter
	- 30 feet high
Media	- 21.5 feet high Surfpac Media
Surface area	- 7.07 sf
Volume	- 153 cf

2 - 0.5 HP pumps for influent and recycle
1 - 1.5 HP pump for dilution

RESULTS

The filter removed less than 60 percent of the BOD under all test conditions (Table II). The highest removal of soluble BOD was 59 percent, achieved with low organic and hydraulic loadings (85 lbs/1,000 cf/day and 407 gpd/sf) with a recycle ratio of 2.5. BOD removals in excess of 45 percent were attained at the same hydraulic load, but with higher organic loadings (282 and 469 lbs/1,000 cf/day) and a recycle ratio of 5.25. In general, BOD removal percentages were reduced as the load increased.

The highest BOD removals per unit volume of media occurred at low levels of dilution with recycle ratios in excess of 5.0. The low dilution levels had a higher temperature which improved treatment efficiency and minimized freezing problems; however, high recycle ratios lowered the temperature. Therefore, if the process were to be used, it would be necessary to achieve a balance between recycling and dilution.

The performance of the pilot plant was characterized by a mathematical expression similar to one proposed by the National Research Council.[4] The expression may be used to predict the effectiveness of the unit under a wide range of operating conditions or applied to full-scale systems with varied loads and recycle rates. The formula was not verified for high levels of treatment as removals of BOD in excess of 60 percent were not attained.

TABLE II

CITY OF KALAMAZOO, MICHIGAN
WASTEWATER TREATMENT PLANT

PLASTIC MEDIA TRICKLING FILTER INVESTIGATIONS

Flow Rate (gpd/sf)	Dilution	Data Points	Recycle Ratio R/Q	Average Temp. (°F)	SBOD Loading* lbs/ 1,000/ cf/day	Influent SBOD (mg/l)	Effluent SBOD (mg/l)	SBOD Removals lbs/ 1,000 cf/day	(%)
407	10:1	6	2.50	--	85.3	543	223	50.3	58.9
611	10:1	10	1.16	--	199.7	848	636	50.0	25.0
611	10:1	18	2.33	51.7	144.1	612	438	41.0	28.4
1426	10:1	8	0.40	53.5	256.6	467	409	31.9	12.4
407	5:1	18	2.50	51.0	141.5	901	561	53.4	37.7
407	2:1	6	2.50	46.8	234.7	1495	918	90.6	38.6
407	2:1	16	5.25	60.0	281.9	1796	983	127.7	45.3
407	1:1	18	5.25	72.5	468.8	2986	1573	221.8	47.3

*Excludes recycle

The performance of the unit at 20°C may be described as follows:

$$E_{SBOD} = \frac{1}{-0.249 + 0.0423\ \sqrt{W/FB}}$$

where

E_{SBOD} = fractional efficiency of soluble BOD removal

W = soluble BOD loading to filter (lbs/day)

V = volume of filter media (acre-ft)

F = recirculation factor = $\frac{1 + R}{(1+R/10)^2}$

R = recirculation ratio

BOD removals during the investigation showed that the process could be used as a roughing device prior to recycling. The major difficulty encountered during this study was the control of foam and odor. Extensive facilities for foam and odor control would be required to develop a viable process.

DISCUSSION

Treatment of wet air oxidation recycle streams using plastic media trickling filters was not deemed to be viable at Kalamazoo. The requirements for dilution water, foam control, and odor control would have resulted in a rather complex treatment scheme with a high potential for creating nuisance conditions. The need to lower the temperature would require either well water or high volumes of primary effluent. The use of clean well water would have increased substantially the flow through subsequent treatment processes. The use of the warm primary effluent would increase pumping requirements to the towers.

The process selected for Kalamazoo was to continue to return the wet air oxidation decant and filtrate directly back to the wet stream treatment train.

CONCLUSIONS

The following conclusions may be derived from the study.

1. Plastic media filters can be used to remove a substantial portion of the soluble BOD in wet air oxidation recycle streams.
2. The process could be used for pretreatment of wastes before they are returned to the wet stream treatment process.
3. The effluent from the process has high BOD and color and is not suitable for direct discharge.
4. The use of plastic media trickling filters would require a significant investment in temperature, foam, and odor control facilities.
5. Treatment of waste oxidation recycle streams using plastic media filters was found to be undesirable due to the high potential for odor, extreme foaming characteristics, and added complexity to the overall waste treatment scheme.

THE LIMA CASE HISTORY

The second case history deals with the use of plastic media trickling filters for nitrification at the City of Lima, Ohio Wastewater Treatment Plant. The pilot studies leading to the design of these facilities have been reported elsewhere[2,3]. The description of the plant and early operating experience also have been reported previously[4]. This paper summarizes the plant design criteria and discusses operating experience for the five years the complete plant has been in operation.

THE PLANT

The improvements to the wet stream facilities were essentially completed in the fall of 1976. Upgrading of the sludge treatment/disposal facilities was completed in mid-1979.

The plant is designed for an average dry weather flow of 18.5 mgd and a peak of 53 mgd. The original design concept called for the secondary and advanced treatment portions of the plant to operate at a peak rate of 33 mgd with the remaining flow receiving primary settling and chlorination.

The upgraded plant includes screening, grit removal, primary settling, aeration, final settling, nitrification towers, chlorination, and phosphorus removal. Ferric chloride and anionic polymer are used for phosphorus removal. Sludge treatment and disposal includes gravity thickening, anaerobic digestion, vacuum filtration, sludge cake storage, and land spreading. Normal sludge treatment/disposal uses thickening, digestion, and land spreading of liquid sludge. Vacuum filtration and sludge storage followed by landfilling is used as backup to land spreading. The design also includes a centralized computer control system. The plant discharges to the Auglaize River.

The design of the two 106 ft. diameter nitrification towers was based on the results of pilot studies. The media for the full-scale facility was supplied by Goodrich. The basis of design, description of the individual treatment units, and projected plant effluent are shown in Table III.

TABLE III

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

DESIGN CRITERIA AND DESCRIPTION OF PLANT

Average Daily Flow: 18.5 mgd
Peak Flow Through Secondary and Tertiary Facilities: 33 mgd
Peak Flow Through Primary Treatment: 53 mgd

Unit	Size	Capacity and/or Operating Conditions
Bar Screens (2)	1 @ 5' wide 1 @ 6' wide	53.0 mgd
Grit Removal Basins (2) 1 @ 24' x 24'	1 @ 20' x 20' 23.1 mgd	16.0 mgd
Primary Settling Tanks (7) 2 @ 3,600 sf 2 @ 4,803 sf 1 @ 4,900 sf	2 @ 2,964 sf	53.0 mgd @ 1,900 gpd/sf 33.0 mgd @ 1,200 gpd/sf 18.5 mgd @ 650 gpd/sf
Aeration Tanks (5)	730,250 cf total	7.1 hrs @ 18.5 mgd 4.0 hrs @ 33.0 mgd
Aeration Blowers (5) 2 @ 10,100 SCFM	3 @ 9,300 SCFM	1,760 cf air/lb BOD applied

TABLE III
(Continued)

Unit	Size	Capacity and/or Operating Conditions
Final Settling Tanks (4)	115' dia. x 14' swd	33.0 mgd @ 794 gpd/sf 18.5 mgd @ 445 gpd/sf
Nitrification Towers (2)	106' dia. x 21.5' deep	18.5 mgd @ 0.73 gpm/sf 18.5 mgd @ 0.18 lbs TK-N/sf 33.0 mgd @ 1.30 gpm/sf 33.0 mgd @ 0.32 lbs TK-N/sf
Chlorine Contact Tanks (2)	37,970 cf total	18.5 mgd @ 24 minutes contact 33.0 mgd @ 15 minutes contact
Phosphorus Removal Chemical Pumps		
$FeCl_3$ (2)	210 gph each	25 mg/l Fe
Polymer (2)	210 gph each	0.2 mg/l
Sludge Thickeners (2)	75' dia. x 11' swd	21.4 lbs/sf primary sludge 2.4 lbs/sf chemical sludge 3.9 lbs/sf secondary sludge
Anaerobic Digesters		
Primary (2)	85' dia. x 22' swd	21 days detention
Secondary (1)	85' dia. x 22' swd	
Sludge Holding Tanks (2)	70' dia. x 32' swd	103,000 cf total

TABLE III
(Continued)

Unit	Size	Capacity and/or Operating Conditions
Vacuum Filters (3)	12' dia. x 10'	1,130 sf total
Supernatant and Filtrate Holding Tank (1)	25' dia. x 8' swd	4,000 cf

Design Effluent Quality

BOD:	9 mg/l @ 18.5 mgd
SS:	14 mg/l @ 18.5 mgd
NH_3-N:	2 mg/l (summer) - 30-day average
	7 mg/l (winter) - 30-day average
P:	1 mg/l - 30-day average

THE NPDES PERMIT

The plant operates under a National Pollutant Discharge Elimination System (NPDES) Permit issued on January 21, 1981 that expires on January 20, 1985. The Permit establishes limitations for three different flow conditions. The pertinent limits of the Permit may be summarized as follows:

	At 18.5 mgd		At 33.0 mgd		At 53.0 mgd	
Parameter	30 Day Avg.	7 Day Avg.	30 Day Avg.	7 day Avg.	30 Day Avg.	7 Day Avg.
Suspended Solids (mg/l)	14	20	20	30	30	--
BOD_5 (mg/l)	9	13	14	19	20	
Fecal Coli-* form/ 100 ml	1,000	2,000	1,000	2,000		
NH_3-N (mg/l) (Summer)	2	4	--	--	--	--
NH_3-N (mg/l) (Winter)	4	8	--	--	--	--
P (mg/l)	1	1.5	--	--	--	--

*Summer Only

The Permit also requires that the D.O. of the effluent be not less than 5 mg/l.

The winter NH_3-N limitation in the Permit is more restrictive than the plant was designed for.

OPERATING CHARACTERISTICS, OPERATION PROBLEMS, AND CORRECTIVE MEASURES

The nitrification facilities were put on line in late summer of 1976. Operation was interrupted shortly afterwards to repair damage to the plastic media in one of the towers. Damage resulted from mechanical failure of one of the distributor arms.

The facility began nitrifying in about eight weeks and was producing the expected effluent values by early November. The time required for the start of nitrification was about the same as in the pilot studies.

Operation of the nitrification facilities has been remarkably free of problems. Early on, the towers operated at almost 100 percent recirculation with no attempts to optimize recirculation rates. For the last couple of years, recirculation has been set at about 11 mgd with the rate varying inversely to flow.

The operational simplicity of the system is greatly appreciated by the plant personnel. The operator simply reviews the computer printouts in the morning and makes any necessary adjustments. The plant personnel claim that operation of the towers and recirculation system "take about a minute a day".

The towers sloughed off solids late in the summer of 1979 for a period of approximately two hours. No decrease in process efficiency was reported following slough-off. No noticeable sloughing of solids has occurred since.

The towers have experienced no significant operating problems during about 5.5 years of operation other than icing two or three times during the winters of 1977 and 1978, two of the coldest winters on record for the area. During these occurrences, ice built up along the filter walls and stopped the distributor arms. The operators broke the ice and the towers were put back into operation.

At the beginning of the winter of 1979, operating personnel capped the end nozzle in each of the distributor arms, eliminating ice formation from splashes on the walls. No icing problems have been experienced since then.

In late October 1981, the bearings in the distributing mechanism in one of the towers failed.

RESULTS

The results for BOD, suspended solids, ammonia, and dissolved oxygen during the first three full years of treatment facilities operation have been reported previously and are summarized in Table IV. The results for 1980 and 1981 are presented in Tables V and VI. Raw wastewater and air temperatures for the same period are shown in Table VII.

The tables show that the quality of the effluent is consistent and generally better than required by the NPDES Permit. The one instance of high NH_3-N in February 1980 has been attributed to analytical error. At the time the plant was experimenting with different analytical methods. The high suspended solids during January, February, March, and April of 1980 were due to the plant's inability to dispose of sludge with the consequent high sludge inventory in the aeration system.

The flows shown in Table V are total flows to the plant and include a small portion of the flow that received primary treatment only during January, February, March, and April of 1980. The actual average flow through secondary treatment and nitrification for these months are:

January	13.14 mgd
February	10.64 mgd
March	22.98 mgd
April	18.80 mgd

Beginning with May of 1980, the computer program governing storage in the sewage system began operating properly. Storage in the system has reduced peak flows to the plant and practically eliminated the need to bypass partially treated wastes.

From 1979 through early 1981, plant personnel took approximately four measurements per month of TK-N in the nitrification towers influent and effluent. The average for these values is shown in Table VIII. Also shown in the table are the loading to the tower (lbs TK-N/sf/day) and the resulting effluent NH_3-N concentration for the years 1979, 1980, and part of 1981.

TABLE IV

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

SUMMARY OF INFLUENT AND EFFLUENT CHARACTERISTICS (1977-1979)

	Flow (mgd)			Average Raw Wastewater			Average Final Effluent				
Year	Avg. Day	High Day	Low Day	BOD (mg/l)	SS (mg/l)	P (mg/l)	BOD (mg/l)	SS (mg/l)	P (mg/l)	NH_3-N (mg/l)	DO (mg/l)
1977	15.8	48.6	8.4	95.4	111.1	9.4	2.3	5.3	1.9	1.4	9.7
1978	12.8	60.8	5.4	102.5	130.5	5.6	3.8	7.0	0.7	1.6	9.9
1979	16.0	48.7	6.4	157.6	126.7	5.1	5.3	9.3	0.8	1.3	11.3

TABLE V

CITY OF LIMA, OHIO

WASTEWATER TREATMENT PLANT
RAW WASTEWATER AND EFFLUENT CHARACTERISTICS (1980)

	Flow (mgd)			Average Raw Wastewater			Average Final Effluent				
Month	Avg. Day	High Day	Low Day	BOD (mg/l)	SS (mg/l)	P (mg/l)	BOD (mg/l)	SS (mg/l)	P (mg/l)	NH_3-N (mg/l)	DO (mg/l)
Jan.	13.77	41.28	8.12	115	210	4.1	4.1	21.9	1.46	0.9	14.1
Feb.	11.98	30.88	6.00	128	118	5.5	4.8	28.5	2.08	5.6	15.3
Mar.	28.22	48.39	10.24	56	128	2.9	13.0	65.9	1.10	3.4	13.1
Apr.	20.92	65.06	10.05	68	123	3.9	7.7	52.8	0.68	0.7	13.5
May	12.50	28.03	7.76	106	97	4.7	5.2	14.3	0.73	1.3	12.7
June	15.37	45.38	6.91	99	131	4.5	2.8	10.7	0.38	1.7	11.9
July	13.10	26.27	7.62	110	161	4.9	5.5	6.8	0.61	1.8	12.1
Aug.	13.33	31.60	9.06	96	125	4.5	2.8	6.9	0.67	0.7	12.0
Sept.	10.77	19.85	7.48	106	114	5.6	1.8	7.2	0.78	0.7	11.5
Oct.	9.99	18.13	6.25	128	155	5.9	3.2	6.9	0.64	0.6	12.8
Nov.	10.21	19.53	7.71	128	112	5.6	5.0	4.7	.75	0.5	14.0
Dec.	10.14	18.10	7.46	132	92	5.3	5.7	3.7	.76	0.8	14.4

TABLE VI

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

RAW WASTEWATER AND EFFLUENT CHARACTERISTICS (1981)

	Flow (mgd)			Average Raw Wastewater			Average Final Effluent				
Month	Avg. Day	High Day	Low Day	BOD (mg/l)	SS (mg/l)	P (mg/l)	BOD (mg/l)	SS (mg/l)	P (mg/l)	NH_3-N (mg/l)	DO (mg/l)
Jan.	9.11	21.44	5.98	111	107	5.2	3.9	4.3	0.74	0.99	13.6
Feb.	14.31	34.10	5.42	87	104	4.3	5.8	7.0	0.84	0.85	12.3
Mar.	8.02	12.50	5.23	104	95	5.7	5.3	6.4	0.94	0.40	13.2
Apr.	16.83	44.33	7.07	73	129	4.4	5.0	11.4	0.55	0.51	11.6
May	19.32	37.58	7.07	58	83	3.2	3.6	8.1	0.43	0.53	12.4
June	16.60	53.54	6.38	59	142	4.5	3.1	20.8	0.50	0.25	10.7
July	9.30	25.91	6.02	92	153	5.4	2.4	6.2	0.63	0.23	10.4
Aug.	8.12	22.17	5.00	115	106	6.0	3.0	6.0	0.67	0.29	10.4
Sept.	10.26	32.01	5.16	98	103	4.8	4.1	5.2	0.58	0.27	10.5
Oct.	9.29	33.09	5.41	117	94	5.5	4.9	5.2	0.59	0.60	10.9
Nov.	9.08	27.53	5.44	85	101	5.4	4.6	4.3	0.67	0.55	10.6
Dec.	12.00	45.40	5.20	87	78	-	4.1	3.3	0.68	0.30	12.0

TABLE VII

CITY OF LIMA, OHIO

WASTEWATER TREATMENT PLANT
RAW WASTEWATER AND AIR TEMPERATURE (1980-1981)

	Avg. Raw Wastewater Temp. (°F)	Air Temp. (°F) Avg. Max.	Avg. Min.
January 1980	47	31	21
February	45	31	17
March	42	44	29
April	49	60	39
May	58	75	51
June	64	79	58
July	70	86	67
August	73	85	68
September	71	79	58
October	63	63	41
November	54	46	32
December	52	33	23
January 1981	48	31	17
February	48	43	24
March	50	48	30
April	56	64	44
May	59	64	48
June	66	78	59
July	72	80	64
August	74	77	58
September	66	68	54
October	60	57	40
November	56	45	33
December	50.9	30.5	21.9

TABLE VIII

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

TK-N LOADING VERSUS NH_3-N IN EFFLUENT (1979, 1980, & 1981)

Month	Tower Influent TK-N (mg/l)	Tower Loading* lbs TK-N/ sf/day	Tower TK-N (mg/l)	Effluent NH_3-N (mg/l)
1979				
January	13.70	0.08	3.48	1.24
February	17.15	0.12	8.43	3.92
March	9.34	0.10	4.18	1.40
April	7.58	0.08	2.30	0.85
May	23.90	0.19	5.59	1.15
June	14.58	0.08	7.92	1.09
July	9.33	0.06	7.78	1.90
August	4.82	0.04	3.81	0.45
September	2.41	0.02	1.68	0.80
October	2.71	0.01	1.62	0.68
November	2.82	0.03	2.10	1.38
December	2.14	0.02	1.97	0.59

TABLE VIII
(Continued)

Month	Tower Influent TK-N (mg/l)	Tower Loading* lbs TK-N/ sf/day	Tower TK-N (mg/l)	Effluent NH_3-N (mg/l)
1980				
January	1.8	0.01	1.5	0.9
February	3.5	0.02	1.0	5.6
March	15.0	0.16	3.8	3.4
April	8.4	0.08	3.7	0.7
May	7.6	0.05	1.8	1.3
June	7.0	0.05	1.5	1.7
July	7.6	0.05	2.4	1.8
August	3.8	0.02	1.0	0.7
September	7.8	0.04	2.7	0.7
October	12.2	0.06	2.4	0.6
November	9.5	0.05	2.4	0.5
December	9.2	0.04	4.0	0.8
1981				
January	9.2	0.04	2.5	1.0
February	11.2	0.08	3.6	0.9

*Total area of towers = 17,650 sf.

SLUDGE PRODUCTION

The nitrification towers are not followed by settling tanks, therefore no sludge is collected. The pilot studies leading to the design showed a sludge collection system would not be required. The findings of the pilot studies have been largely confirmed during operation. The towers have sloughed off solids noticably only once for a period of about two hours during about five and one-half years of operation.

IMPACT ON RIVER WATER QUALITY

Table IX shows the BOD and D.O. concentrations measured upstream and downstream from the plant in 1980 and 1981. Downstream measurements are taken approximately 400 yards below the plant discharge. It is evident from the table that the plant effluent has essentially no impact on the Auglaize River water quality.

OPERATION AND MAINTENANCE COST

Operation and maintenance costs averaged $142.55 per million gallons in 1978, $138.93 per million gallons in 1979, $178.50 in 1980, and $244.82 in 1981. Itemized operation and maintenance costs for the last two years are shown in Table X.

The plant operator reports that practically no manpower is required to operate the nitrification towers and appurtenant pumping station. For all practical purposes power and maintenance are the only operating expenses associated with the towers.

TABLE IX

CITY OF LIMA, OHIO
WASTWATER TREATMENT PLANT
UPSTREAM AND DOWNSTREAM AUGLAIZE RIVER WATER QUALITY* (1980-81)

	Upstream		Downstream	
	BOD_5 mg/l	D.O. mg/l	BOD_5 mg/l	D.O. mg/l
January 1980	4.4	16.1	4.5	15.8
February	4.6	19.7	3.4	17.5
March	4.4	14.2	4.8	14.1
April	3.1	14.1	3.6	13.7
May	6.7	13.1	5.9	12.8
June	4.7	11.1	4.0	10.9
July	6.8	11.2	4.1	10.9
August	5.0	10.3	4.1	10.7
September	5.6	9.9	3.1	10.2
October	5.8	11.8	3.6	12.4
November	4.7	13.7	4.1	13.2
December	4.4	15.5	4.7	14.4
January 1981	3.6	15.0	4.5	14.1
February	7.2	13.7	6.5	13.5
March	4.4	16.6	3.8	16.1
April	4.3	13.3	5.3	12.5
May	5.5	11.9	4.5	11.2
June	2.5	8.7	3.7	9.4

TABLE IX
(Continued)

	Upstream		Downstream	
	BOD_5 mg/l	D.O. mg/l	BOD_5 mg/l	D.O. mg/l
July	4.4	10.8	3.4	10.3
August	9.1	14.1	5.2	12.8
September	7.0	7.4	4.5	9.9
October	6.6	10.5	5.3	10.4
November	5.5	13.9	5.5	13.0
December	5.1	13.9	5.6	13.4

* Values in Table represent daily samples through January 1981 and once per week samples beginning with February 1981.

TABLE X

CITY OF LIMA, OHIO
WASTEWATER TREATMENT PLANT

OPERATION AND MAINTENANCE COST (1980 AND 1981)

Item	1980	1981
Payroll	$460,590.42	$ 510,036.08
Power	196,546.12*	191,114.18
Chlorine	5,208.38	3,771.59
Chemicals**	102,870.05	179,192.08
Miscellaneous	162,265.17	172,260.34
	$927,480.14	$1,056,374.20

* The cost for power averaged 0.022/KWH.
** Ferric chloride and polymer.

DISCUSSION

Figures 2 and 3 show the data derived from the pilot studies and the operating results for 1979, 1980, and part of 1981; the only years that the plant has collected data regularly on the TK-N concentration in the activated sludge effluent. The results predicted by the pilot studies have been confirmed under actual operation. This strongly supports the concept of designing nitrification towers on the basis of TK-N loads.

The nitrification efficiency of the total system meets design expectations. For part of the year, the secondary plant nitrifies well, as evidenced by the low TK-N in the secondary effluent. During that time, the towers function as polishing facilities. The towers always oxidize substantial amounts of TK-N. The additional TK-N oxidation in the towers results in a very stable effluent with very low TK-N values. For the twelve-month period of March 80 through February 81, the TK-N in the effluent averaged only 2.65 mg/l. The organic Nitrogen in the effluent averaged 1.47 mg/l.

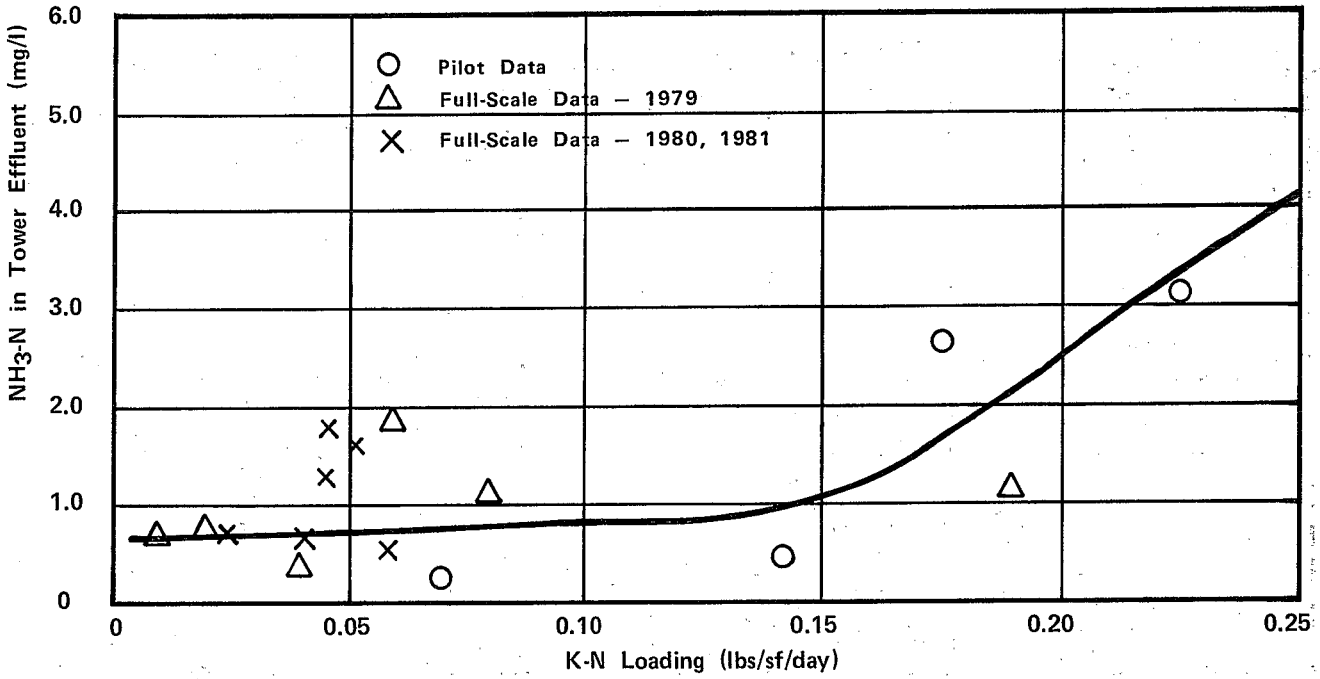

FIGURE 2

K-N LOADING VS. NH_3-N IN TOWER EFFLUENT – SUMMER DATA

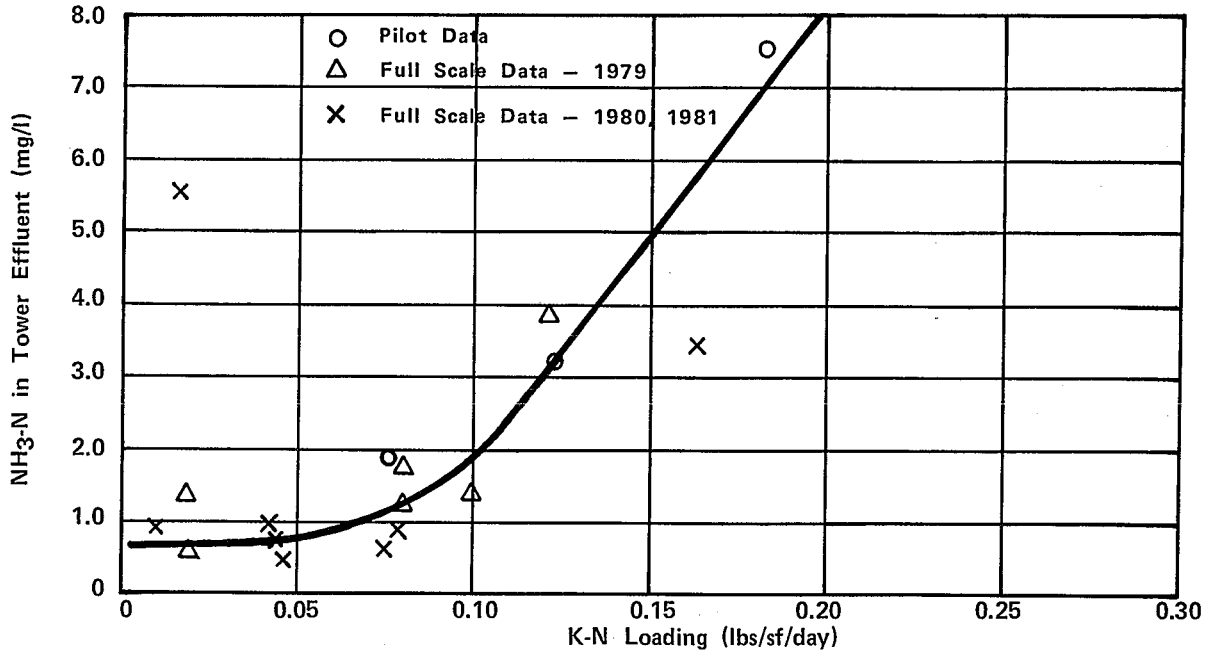

FIGURE 3

K-N LOADING VS. NH_3-N IN TOWER EFFLUENT – WINTER DATA

The tower effluent is always better than the influent. Even though substantial nitrification occurs in secondary treatment, the towers are needed to consistently meet NH_3-N effluent requirements.

The towers are not generally loaded to their design capacity. Plots of TK-N applied (lbs/sf/day) vs. TK-N oxidized (lbs/sf/day) show a straight line relationship indicating that loadings have not been the limiting factor in TK-N oxidation.

During the summer months, the NH_3-N concentration in the plant effluent is about the same as for loadings ranging from 0.01 to 0.15 lbs TK-N/sf/day. During the winter months, and over the same loading range, the effluent NH_3-N concentration increases substantially as the TK-N load increases.

The plant has produced the desired results while treating the highly variable wastewaters generated by a partially combined sewerage system. In any single month, the flow can range from one third to almost three times the design average. The monthly average for BOD in the raw sewage has ranged from 53 mg/l to 246 mg/l. The monthly average for suspended solids in the raw sewage has ranged from 83 mg/l to 210 mg/l. It is unlikely that single or two-stage activated sludge systems could provide as reliable a treatment level under such adverse conditions.

Stable performance is achieved with a minimum of operational adjustment to the nitrification facility. During 1981, the operators simply kept the recycle flow at 11 mgd.

CONCLUSIONS

The following conclusions may be derived from the first five years of operation of the activated sludge/nitrification tower process used at Lima:

1. Nitrification towers following activated sludge consistently produce a high quality effluent with low BOD, suspended solids, TK-N, and NH_3-N, and very high dissolved oxygen concentration.

2. The treatment process is extremely reliable and able to withstand shocks.
3. The plant effluent is normally saturated or super saturated with oxygen. The combination of high D.O. and BOD results in an effluent that exerts little or no oxygen demand on the receiving stream.
4. The towers are easy to operate.
5. The performance of the full-scale facility has confirmed the design criteria derived from pilot studies. Very low ammonia concentrations can be obtained even during the cold winters experienced in midwestern United States.
6. Final settling following the towers has not been necessary as the effluent contains very low suspended solids.

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PART VII: NITRIFICATION AND DENITRIFICATION

NITRIFICATION OF A MUNICIPAL TRICKLING FILTER EFFLUENT USING ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

This paper presents the results of a pilot plant investigation and design for achieving seasonal nitrification of a secondary trickling filter effluent. The municipal treatment plant in Milford, Massachusetts discharges into the headwaters of the Charles River, a Class B stream. Consequently, the Commonwealth of Massachusetts has required a nitrified effluent, phosphorus removal, and effluent filtration. Table 1 lists the present and proposed discharge limitations formulated by the U.S.E.P.A. and the Massachusetts Division of Water Pollution Control.

Table 1. Present Discharge Limitations

Effluent Characteristic	Monthly Average	Maximum Day
Flow (MGD)	4.0	
Biochemical Oxygen Demand 5-day 20° C	30 mg/1	50 mg/1
Total suspended solids	30 mg/1	50 mg/1
Settleable Solids		0.3 ml/1
Fecal Coliform Bacteria	200/100 ml	400/100 ml
Total Coliform Bacteria	1000/100 ml	2000/100 ml
pH	Shall remain between 6.0 and 9.0	

Proposed Discharge Limitations

Effluent Characteristic	Monthly Average	Maximum Day
Flow	4.3	
Biochemical Oxygen Demand 5-day 20° C	7 mg/1	11 mg/1
Total suspended solids	7 mg/1	11 mg/1
Settleable Solids	0.1 ml/1	0.3 ml/1
Total Ammonia Nitrogen	1.0 mg/1	1.5 mg/1
Total Phosphorus	1.0 mg/1	1.5 mg/1
Dissolved Oxygen	-	Not less than 6 mg/1
Fecal Coliform Bacteria	200/100 ml	400/100 ml
pH	Shall remain between 6.0 and 9.0	

After preliminary evaluation of alternate methods of achieving nitrification, rotating biological contactors were selected. A pilot plant study was conducted during the late summer and fall of 1977 to determine both the performance of such a system in nitrifying a trickling filter effluent and the design and operating parameters. Specifically the objectives of the pilot plant operation were to determine various design and operational parameters of the RBC process

with regard to nitrification, including loading rates, removal rates, removal efficiencies, solids production, and solids settling characteristics.

Existing Wastewater Treatment Plant

The existing sewage treatment plant consists of primary sedimentation, raw sludge dewatering, secondary treatment by high rate trickling filters with a varying degree of recirculation, final clarification, and chlorination of the effluent before discharge to the Charles River. Flow enters the plant through two pipe lines, each equipped with a Parshall flume and its attendent flow recording equipment. One line receives the discharge from the Charles Street pumping station through variable speed pumping equipment, and that discharge is therefore responsive to inflow variations. The other line is the main gravity flow outfall from that portion of the collection system not draining to the Charles Street pumping station, and is equipped with a comminutor in addition to the flow meter.

Grit removal is accomplished in the primary settling tanks along with scum and settled sludge removal. The combined grit and raw sludge are then dewatered by vacuum filtration, with the aid of lime and ferric chloride as sludge conditioners. Dried sludge, with a pH of 11 to 12, is disposed of as land fill in the discontinued sand filter beds. The effluent from the primary settling tanks flows through a flow control chamber to 3 trickling filter rotary distributors which uniformly disperse the wastewater onto stone media filters.

Trickling filter effluent is pumped to the final clarifier and a proportionate amount of this effluent is recirculated to the trickling filter inflow, in order to provide the hydraulic quantity necessary for proper operation of the filter distributors during periods of low flow. Effluent from the final clarifier is subsequently chlorinated in a chlorine contact chamber before discharge.

The wastewater flow at Milford is predominately domestic in character, and presently contains very little so-called "industrial" wastewater. Analyses of the raw wastewater reinforce this conviction. It is anticipated that this situation will not substantially change in the future. Furthermore, any future industrial wastes, which might affect proper treatment plant operation will be required to receive adequate pretreatment at the point of origin prior to their acceptance by the municipal treatment facility.

Plant records evaluated over the past five years indicate the average daily flow to the treatment facility is 2.4 M.G.D. and the average influent BOD-5 is 140 mg/1. Effluent BOD-5 averages 25 mg/1., which corresponds to a BOD-5 removal rate on the order of 82%. The facility serves a population of approximately 22,500.

Pilot Plant Description

During the late summer and fall of 1977 a pilot plant was operated on the Milford Wastewater Treatment Plant site. Figure 1, illustrates the flow scheme used for this pilot plant operation.

As shown in Figure 1, the clarified trickling filter effluent was pumped from the effluent trough of the clarifier to the two RBC units by individual submersible pumps. The rate of flow to each RBC unit was controlled by a valve following each pump.

The RBC unit consisted of a semi-cylindrical tank 4.23 feet long and 4.0 feet in diameter which was divided into four equal-volume compartments by means of 1/4-inch steel partitions. Twelve 47-inch diameter rotating polystyrene discs per compartment were mounted on a center drive shaft so that approximately 29% of the disc surface area was submerged in the tank contents. Each unit contained a total of 1570 square feet of surface area and a net tank volume of 120 gallons with the discs in the tank.

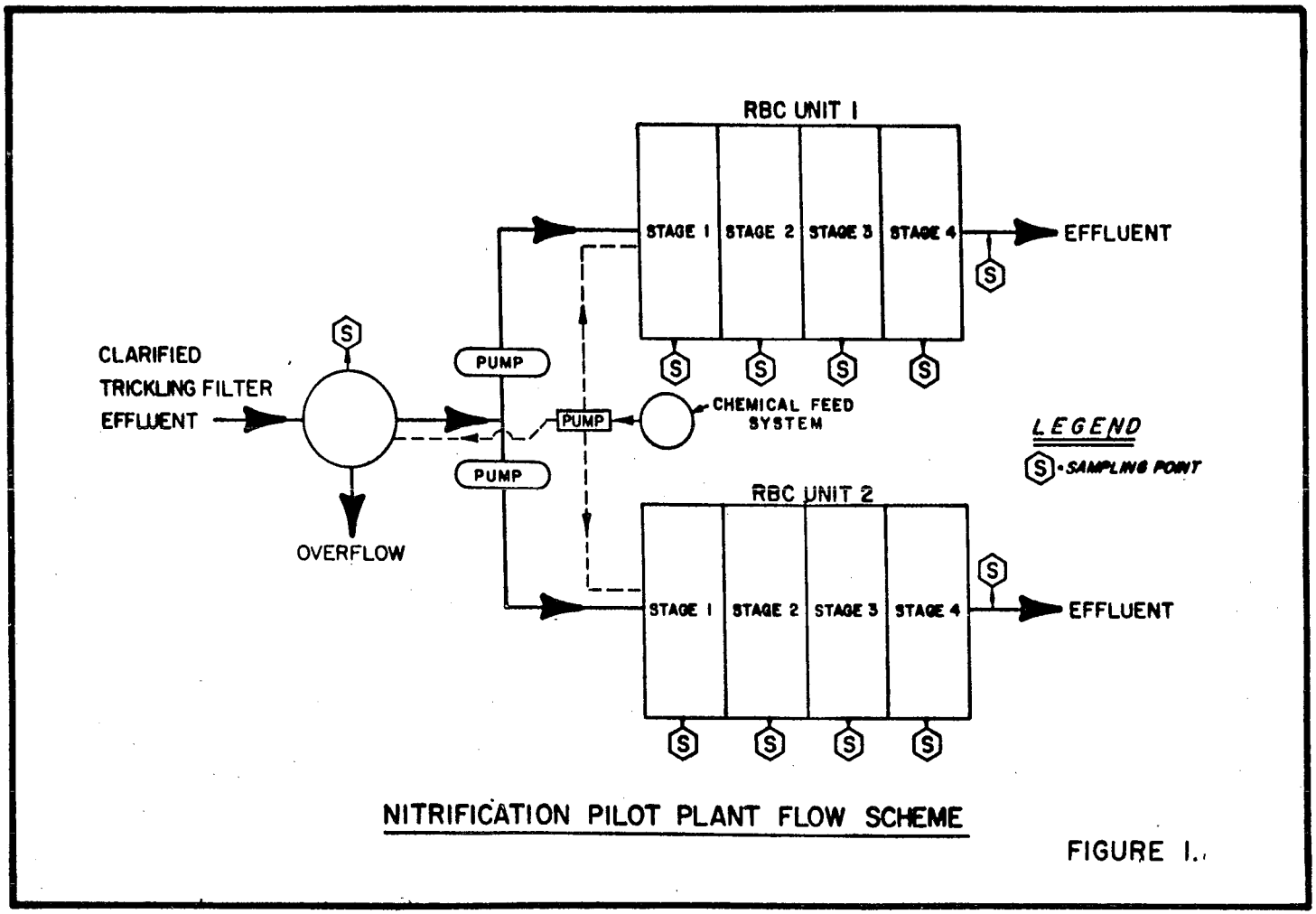
Connections between the compartments or stages permitted four-stage series operation. These connections were made by means of external pipes which by-passed the compartment partitions on the outside of the tank. The wastewater therefore flowed through the unit perpendicular to the center shaft.

Alkalinity additions were made by means of peristaltic pumps from two 55-gallon tanks to the first stage of each RBC unit. The chemical solution was kept mixed by means of a submersible pump in the bottom of the tank.

The effluent from the RBC unit then flowed by gravity to a settling tank, where solids were permitted to settle out and the overflow was discharged to the control chamber adjacent to the existing secondary clarifier.

Pilot Plant Operation

After a two week startup period in the beginning of August 1977, the pilot plant acquired a reasonably full growth of biomass. From that time until December 1977



NITRIFICATION PILOT PLANT FLOW SCHEME

FIGURE 1.

performance data was measured.

It has been stated in the literature concerning nitrification that any biological process for the conversion of ammonia to nitrate will require a source of alkalinity in the amount of 7.14 mg/l of $CaCO_3$ for each 1.0 mg/l of NH_3-N oxidized. (1, 2, 3). The RBC process is no exception and since the influent to the RBC's contained an average of only 51 mg/l as $CaCO_3$, an external source of alkalinity was necessary to insure complete nitrification of the ammonia present.

Alkalinity was supplied to the units in the form of either bicarbonate of soda, soda ash or lime. Efforts were made during the pilot plant study to maintain a pH of 7.0 to 8.5 and an alkalinity of greater than 200 mg/l.

Sampling was arranged so that observations could be made under various conditions and at different times during the day. A 24-hour sampler was utilized a number of times to monitor the characteristics of the influent to the RBC units during the course of a day. In addition, the regular sampling procedure for monitoring the performance of the units was arranged to provide data of both morning and afternoon operation, because the lowest influent concentrations of NH_3-N were observed between 5 A.M. and 10 A.M. and the highest between 3 P.M. and midnight.

Operational Results

Table 2 illustrates the typical unchlorinated trickling filter effluent for summer operation which served as the influent to the RBC pilot plant.

Table 2. Typical Unchlorinated Trickling Filter Effluent

Parameter	Typical Value
TKN, mg/l	10
NH_3-N, mg/l	6
NO_3-N, mg/l	9
BOD_5, mg/l	25
Suspended Solids, mg/l	47
Volatile Suspended Solids, mg/l	38

Parameter	Typical Value
pH *	6.0 to 7.0
Alkalinity (as $CaCO_3$), mg/l *	51

* Values are before adjustment by chemical addition for pilot plant operation.

Due to variations in the strength of the raw wastewater entering the plant there was variability in the trickling filter effluent quality. Table 3 depicts the variation in ammonia nitrogen concentration with time for three twenty-four hour periods.

Table 3. RBC Influent Concentration Variation Over a 24 hour Period

Aug. 11-12 Time	NH_3 mg/l	Aug. 14-15 Time	NH_3 mg/l	Aug. 16-17 Time	NH_3 mg/l
11:30 AM	8.4	4:30 P.M.	8.8	7:30 AM	4.7
12:30 PM	-	5:30	12.0	8:30	
1:30	9.8	6:30	9.4	9:30	5.8
2:30	-	7:30	11.3	10:30	
3:30	11.0	8:30	-	11:30 AM	8.4
4:30	8.8	9:30	7.7	12:30 PM	
5:30	8.4	10:30	10.0	1:30	11.5
6:30	-	11:30 PM	8.3	2:30	
7:30	-	12:30 AM	8.9	3:30	12.9
8:30	7.8	1:30	8.9	4:30	
9:30	7.7	2:30	7.4	5:30	11.2
10:30	11.3	3:30	9.8	6:30	
11:30 PM	9.5	4:30	8.2	7:30	10.0
12:30 AM	12.0	5:30	7.4	8:30	
1:30	10.0	6:30	7.0	9:30	11.2
2:30	9.4	7:30	6.9	10:30	
3:30	9.4	8:30	6.8	11:30 PM	7.4
4:30	9.2	9:30	6.3	12:30 AM	
5:30	8.0	10:30	6.6	1:30	7.0
6:30	6.5	11:30 AM	-	2:30	
7:30	6.6	12:30 PM	11.0	3:30	10.0
8:30	7.0			4:30	
9:30	10.2			5:30	6.0

Note: For most loadings NH_3 Ammonia Range 5 to 13 mg/l

Analysis of the performance of the pilot plant was begun after an adequate biomass had been established on the surface of the media. From this point and throughout the course of the study it was apparent that the RBC units could consistently produce an effluent with less than 0.5 mg/l of ammonia, providing that sufficient alkalinity was available.

The establishment of loading criteria therefore remained as the primary concern of the study. Figure 2 illustrates the results of the analyses performed from mid-August through November, 1977, and indicates a linear relationship between pounds of ammonia applied to the RBC units and pounds of ammonia removed, for the range 0.04 to 0.4 pounds NH_3-N/1000 ft^2-day. This linear relationship between pounds applied and pounds removed has been established in other studies. (3) (4).

Figures 3 and 4 present the efficiency of the RBC units with respect to ammonia removal. Figure 3 represents all the data points while Figure 4 contains only those points for which the influent ammonia concentration to the RBC units was greater than or equal to 6.0 mg/l. It is evident from these efficiency plots that the removal of ammonia was greater than 95 percent for loadings up to 0.2 pounds NH_3-N applied per 1000 square feet of surface area-day.

In addition, for loadings up to 0.4 pounds NH_3-N applied 1000 s.f.-day, the removal rate was generally greater than 90 percent.

During the course of the pilot study the temperature ranged from 9^{o} C to 19.5^{o} C with the typical value being 16^{o} to 17^{o} C.

In this range, the temperature did not appear to have a significant affect on the efficiency of ammonia removal in the RBC units.

During the course of the study a number of sequential sampling runs were conducted following an immediate increase in the hydraulic load to simulate a peak flow. The sequential sampling was done in an attempt to trace a plug of wastewater flow through the unit.

Figures 5 through 7 show the ammonia nitrogen concentration levels through the pilot units for both pre-peak equilibrium loading rates and peak loading rates. Table 4 indicates the ammonia loading rates for the same peaking experiments.

At application rates lower than 0.2 lb NH_3-N/day-1000 s.f. the amount of ammonia removed virtually doubled as the application rate doubled. As indicated in Figures 5 thru 7

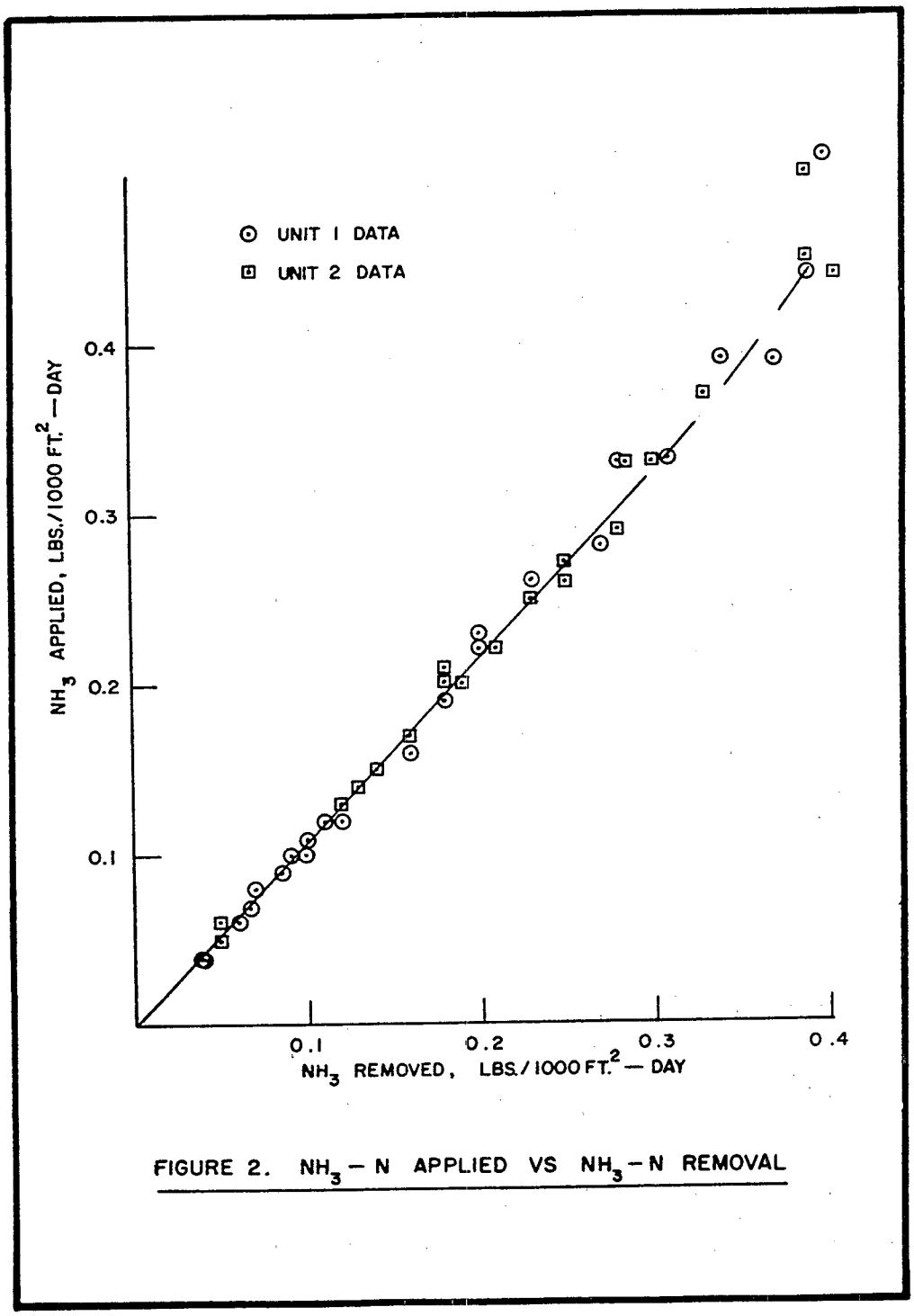


FIGURE 2. NH_3 – N APPLIED VS NH_3 – N REMOVAL

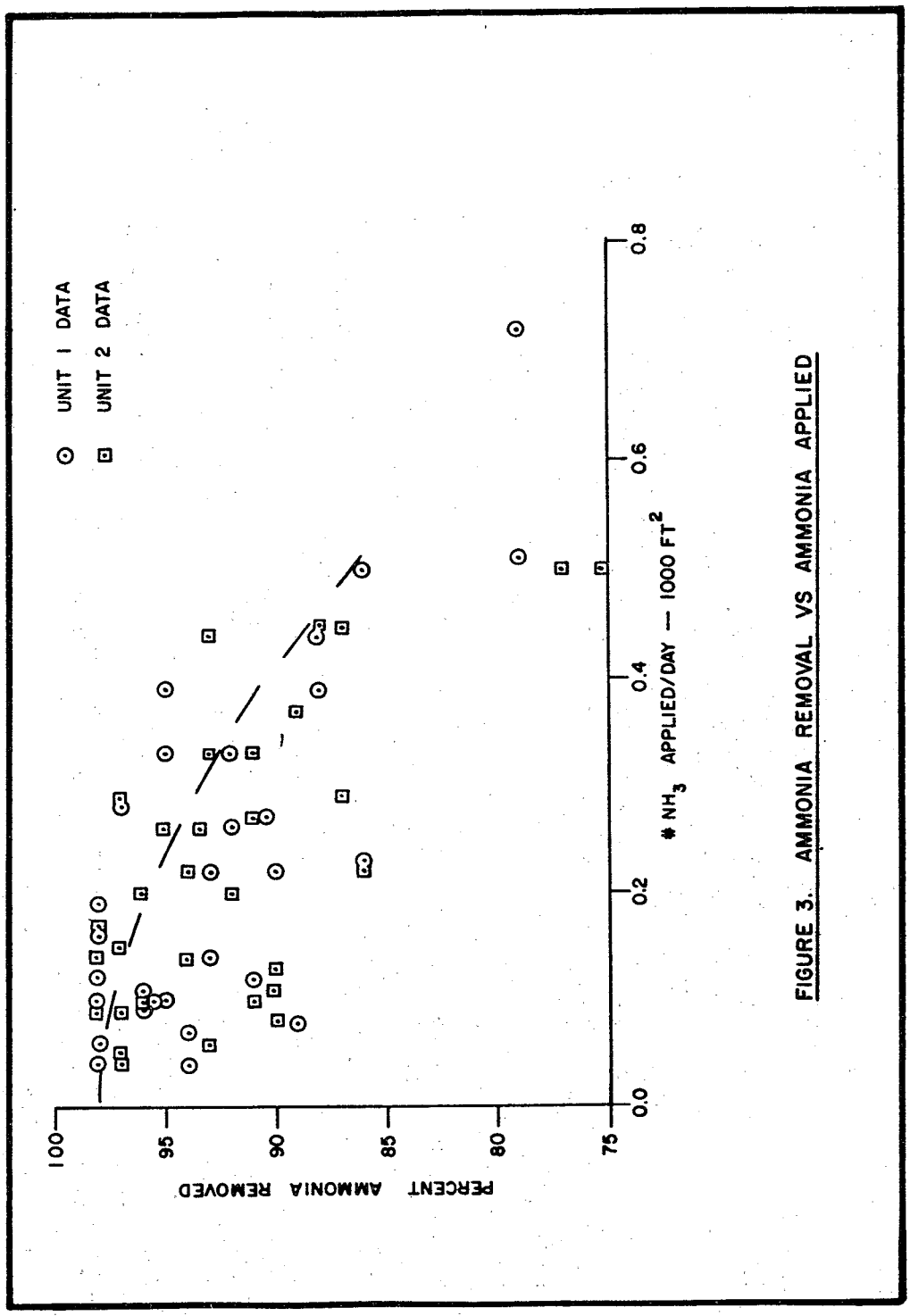


FIGURE 3. AMMONIA REMOVAL VS AMMONIA APPLIED

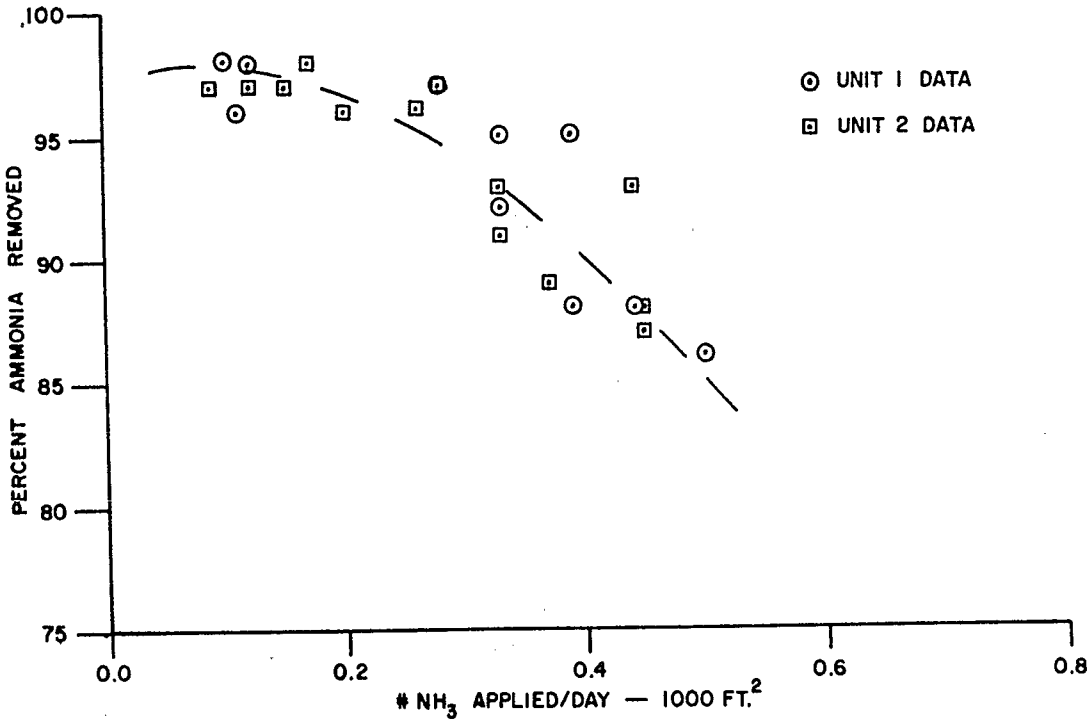
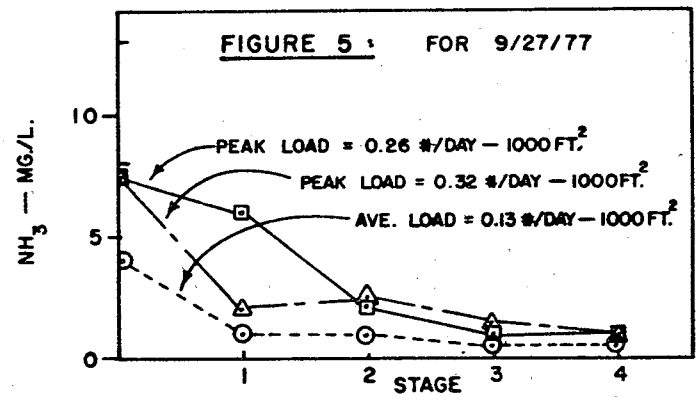
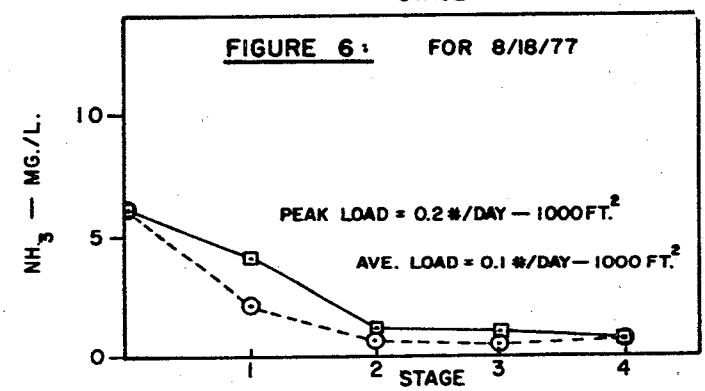
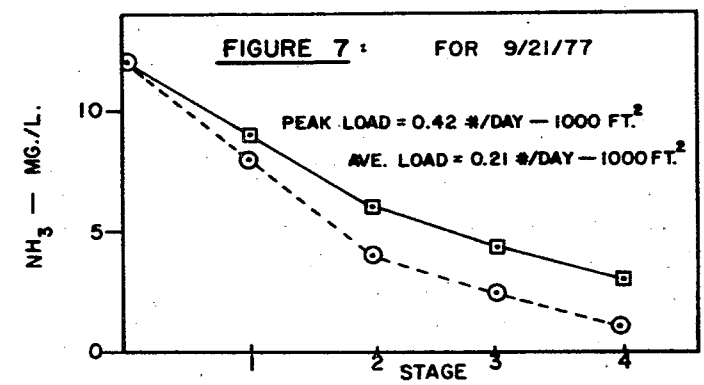


FIGURE 4. PERCENT AMMONIA REMOVAL VS APPLIED AMMONIA FOR INFLUENT VALUES 6.0 MG. NH_3 — N/L AND GREATER. (EXCLUDING PEAK LOADINGS)







PEAK LOADING TEST RESULTS

the effluent concentrations of ammonia were slightly higher under peak conditions. With the exception of the run when the peak loading factor was increased from 0.2 to 0.4 lb-NH_3-N per day per 1000 $ft.^2$ shown in Figure 7, the effect of 4 stage operation attenuated decreases in removal efficiency due to peaking. This was the only case where the effluent ammonia concentration exceeded 1 mg/l.

Table 4. Effects of Peak Loading on NH_3-N Removal

Q (gpm)	Detention Time (hrs)	NH_3-N Applied #/day-1000 $ft.^2$	Removed #/day-1000 $ft.^2$	% Removal
0.8	2.5	0.1	0.098	-
1.6 *	1.25	0.22	0.20	91
1.1	1.82	0.16	0.155	-
2.2 *	0.91	0.33	0.306	93
1.3	1.54	0.13	0.125	-
2.6 *	0.77	0.25	0.23	92
1.3	1.54	0.1	0.95	-
1.85 *	1.08	0.26	0.239	92
1.3	1.54	0.13	0.126	-
2.25 *	0.88	0.32	0.286	89
0.8	2.5	0.1	0.98	-
1.6 *	1.25	0.22	0.187	85
0.9	2.22	0.2	0.187	-
1.85 *	1.08	0.42	0.320	76
1.0	2.0	0.11	0.108	-
1.9 *	1.05	0.21	0.192	92

* Flow values from an instant flow peak and represent the loading which gave the least efficient treatment.

Table 4, which summarizes the significant parameters observed during the peak loading studies, indicates that good removal efficiencies were maintained even when substrate loading rates were doubled and hydraulic retention time halved.

The results of the pilot study indicated that the effluent BOD_5 from the RBC units averaged approximately 8 mg/l.

These samples were either collected from the 55 gal. drum clarifiers, or samples collected from stage 4 of the RBC's and allowed to settle for 30 minutes. The soluble portion of the effluent was approximately 56% and averaged 4 to 5 mg/l of BOD_5. The addition of alum to the last stage of the pilot RBC units further reduced the effluent five day BOD to an average value of less than 2 mg/l.

The average suspended solids content of the influent was 47 mg/l, the fourth stage of the RBC contained about 53 mg/l, whereas the effluent from the units averaged 32 mg/l., of which approximately 80 percent were volatile.

For hydraulic loadings up to 1.80 gpm/s.f. (2.0 gpm) there appeared to be little to no net solids production. The settled effluent averaged 16 mg/l of suspended solids and it would appear that the suspended solids in the effluent from a clarifier following the RBC can be maintained at under 20 to 30 mg/l for low hydraulic loading rates. The solids determinations during the pilot studies included suspended solids from the existing trickling filters which contributed a relatively high percentage of colloidal matter.

Processes Selection and Design of the Wastewater Treatment Facility

Four processes for the removal or conversion of ammonia were investigated to determine their cost-effectiveness, in accordance with EPA criteria. The four processes were as follows: (a) nitrification using Rotating Biological Contactors (RBC); (b) nitrification by aeration in a plug flow reactor; (c) ammonia removal by breakpoint chlorination; and (d) ammonia removal by selective ion-exchange.

Nitrification using RBC was selected as the lowest total life-cycle cost process, although its capital costs are somewhat higher than the other processes. In addition, the overall advantages of process stability and flexibility as well as substantially lower operation and maintenance other than occasional lubrication of the drives and it requires significantly less attention during operation than does the aeration system.

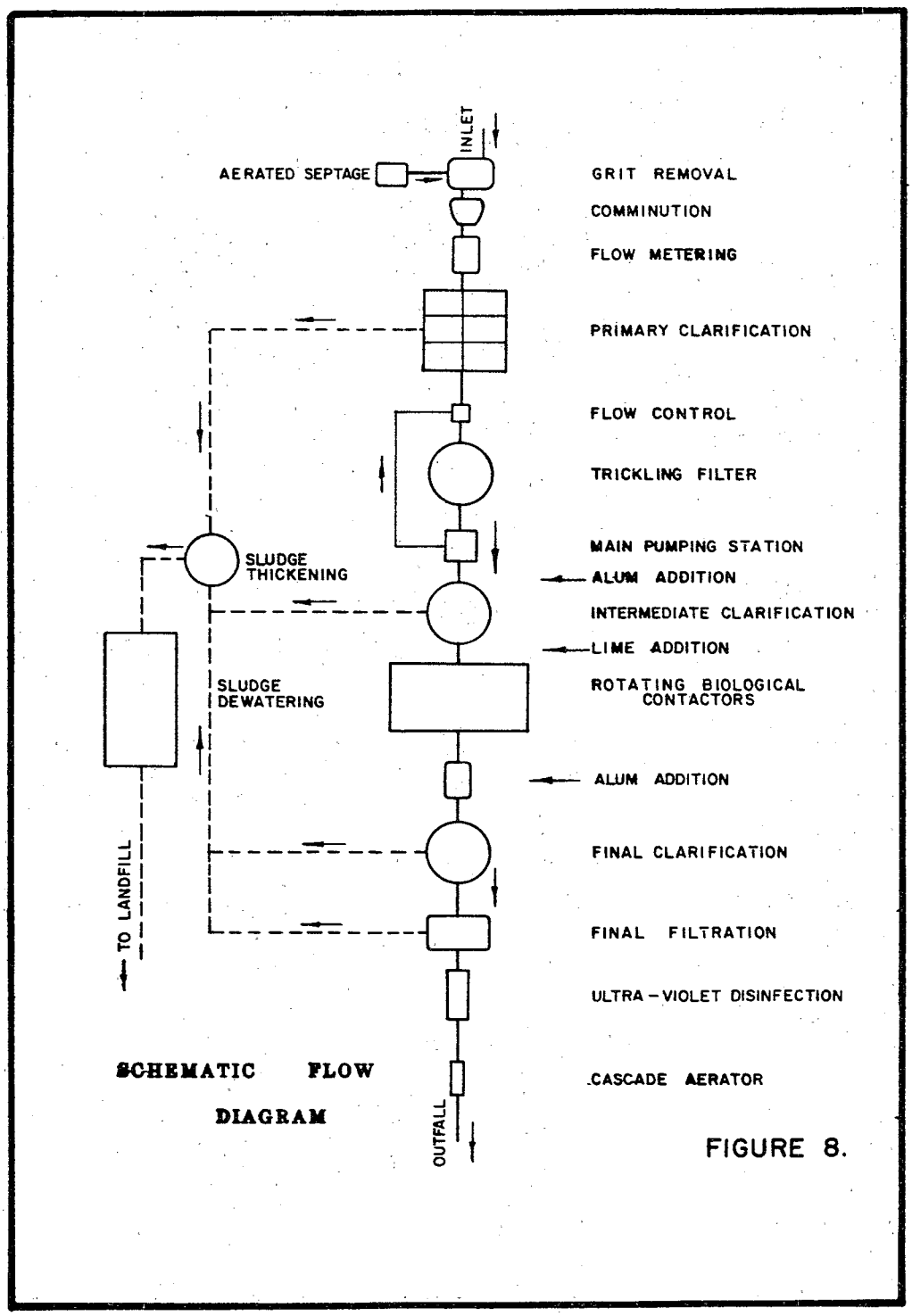
In order to utilize the existing facilities to the most economical degree, and to keep the operation and maintenance costs as low as practical, the present method of treatment by primary sedimentation followed by trickling filters will be retained. As shown previously, the performance of the plant as of late has been of the highest quality to be

expected of that type of process. In addition, the plant personnel is thoroughly familiar with the operation of the existing treatment facility and could therefore maintain the present standards of quality with little difficulty.

The proposed additions to the plant will provide tertiary treatment to reduce the levels of phosphorus and ammonia-nitrogen in the effluent. This secondary stage in the treatment process ideally complements the trickling filter operation because trickling filters will produce an effluent with sufficient BOD to maintain the nitrifying organisms (See Figure 8).

The pilot plant study established that the RBC process is capable of achieving greater than 95 percent removal of ammonia up to an ammonia loading rate of 0.2 pounds NH_3-N per 1000 square feet of surface area per day, provided that sufficient alkalinity is available in the wastewater. At the projected ammonia load for the year 2000, some 3,200,000 square feet of surface area would be required for this loading rate. In order to provide this amount of surface area, it was decided that four (4) trains consisting of six (6) shafts, each 25 feet long, would be utilized. It was felt that the four separate trains would provide the desired degree of flexibility in the system and six shafts per train would insure that sufficient surface area was provided to prevent ammonia breakthrough due to peak flow loadings. The first two (2) shafts of media in each train will be fabricated of so-called "standard density" media, that is, approximately 100,000 s.f. per shaft. The remaining four (4) shafts in each train will be fabricated of "high density" media or about 150,000 s.f. per shaft. In this manner, problems associated with "bridging" of the media due to overloading in the first and second stages can be avoided. In addition, the six stages of media provide sufficient surface area for carbonaceous BOD removal in the first two stages of each unit.

Flexibility in each train will be provided through the utilization of removable wooden baffles between shafts. The amount of surface area per stage and the number of stages per train can then be varied to suit operating conditions. Alkalinity will be provided immediately before the RBC process in the form of hydrated lime. Hydrated lime was chosen as the source of alkalinity due to its lower cost and local availability. It is anticipated that the alkalinity will be a minimum of 200 mg/l in the influent to the RBC.



SCHEMATIC FLOW DIAGRAM

FIGURE 8.

Flexibility was also designed into the phosphorus removal system, in that two different chemicals for the removal of phosphorus can be stored and fed at one time, and chemicals can be added and mixed with the wastewater at three separate locations. Laboratory treatability studies established that alum would provide sufficient removal of phosphorus to meet the discharge limitation of 1.0 mg/l consistently. In addition, the study established that the addition of alum immediately following the trickling filters would be the least costly method. This is due to the low pH (5.8 to 6.2) of the trickling filter effluent. Removal of phosphorus with alum after the RBC process would require substantially more chemical due to the elevated pH of the wastewater at this point. However, chemical addition and mixing facilities will be provided after the RBC process for removal of phosphorus in the final clarifiers. In addition, chemical additions can be made to the headworks at the aerated grit chamber for phosphorus removal in the primary clarifiers.

SUMMARY AND CONCLUSIONS

1. A linear relationship between pounds of ammonia applied versus pounds of ammonia removed existed for ammonia loadings ranging from 0.04 to 0.4 pounds NH_3-N per 1000 s.f.-day.
2. That the removal of ammonia was greater than 95 percent for loadings up to 0.2 pounds NH_3-N applied/1000 s.f.-day, and generally above 90 percent removal for loadings up to 0.4 pounds NH_3-N applied/1000 s.f.-day.
3. That the RBC units were able to respond well to peak hydraulic loading.
4. That there is little to no net production of solids within the RBC units. The solids which were produced did not exhibit good settling characteristics.

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IMPROVEMENT OF NITRIFICATION IN ROTATING BIOLOGICAL CONTACTORS BY MEANS OF ALKALINE CHEMICAL ADDITION

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INTRODUCTION

The need to achieve compliance with ammonia-nitrogen discharge limitations and the current emphasis on energy conservation have resulted in the utilization of RBC technology for the nitrification of secondary wastewater effluents. By the end of the 1970s, four pilot scale efforts, independent of the RBC industry, had been completed which demonstrated that the RBC could nitrify successfully secondary wastewater effluent (1, 2, 3, 4). In 1979, approximately 70 percent of the RBC systems in the United

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States were designed to remove carbonaceous biochemical oxygen demand (CBOD). Another 25 percent of the RBC systems were designed to remove CBOD and for nitrification in the same RBC units. The remaining 5 percent were constructed to nitrify secondary wastewater effluents in order to achieve ammonia-nitrogen effluent discharge limitations (5) Initial evaluations of full scale nitrifying RBC facilities reveal that they have not been completely satisfactory (6,7) Hitdlebaugh (7) reported that an RBC facility, built for CBOD removal and nitrification, failed to meet design specifications during both winter and summer operations. The inability to meet CBOD and ammonia-nitrogen limitations during the summer was attributed to relatively low dissolved oxygen (DO) concentrations (less than 1 mg/1) in the initial nitrifying stages and a low pH (less than ph 7.0) in the latter nitrifying stages. The DO level increased during winter operations and CBOD was removed sufficiently to achieve design expectations. Ammonia-nitrogen removal also improved during the winter but not sufficiently to achieve design projections or effluent limitations. Recommendations from this study included the use of alkaline chemical feed systems to maintain optimum pH levels in order to improve nitrification.

Nitrification within the RBC biofilm is essentially a two-step microbiological process which utilizes two groups of autotrophic bacteria of the family Nitrobacteraceae. The first group of bacteria oxidizes ammonia to nitrite and the second group of bacteria oxidizes nitrite to nitrate. The Nitrosomonas and Nitrobacter genera are considered to be the predominant nitrifying bacteria inhabiting the wastewater environment. Heterotrophic nitrification also occurs when nitrite or nitrate is produced from organic or inorganic compounds by heterotrophic organisms. Over 100 heterotrophic species (including fungi) have been identified which are capable of heterotrophic nitrification. However, the overall contribution to the oxidized nitrogen forms by heterotrophic nitrification is considered to be relatively small (8).

The growth rates of nitrifying bacteria are much slower than the growth rates of heterotrophic bacteria. This important distinction accounts for the inability of nitrification to proceed simultaneously with CBOD removal when high concentrations of organic material (greater than 30 mg/1 of BOD) are present in the wastewater. Minimum doubling times reported for the ammonia-oxidizing bacteria are from 8 to 17 hours (9). Because the growth rates for nitrite-

oxidizers are greater than the growth rates for ammonia-oxidizers (9,10), elevated nitrite concentrations normally do not persist and the ammonia-oxidation step controls the total amount of ammonia which is oxidized to nitrate within the wastewater environment. Carbon dioxide (CO_2) is the carbon source for these autotrophic nitrifying bacteria (11). Although some nitrifying bacteria have been observed to use organic compounds, they were not observed to utilize these organic compounds as the sole carbon source for growth (12). The generation of bacterial biomass per unit of ammonia oxidized (cell yield) is quite small. The total yield for both Nitrosomonas and Nitrobacter has been observed to be from 0.06 to 0.20 gram of cells per gram of ammonia oxidized (9). The nitrification of 20 mg/l of ammonia-nitrogen generates approximately 2 mg/l of solids (10). Therefore, the net amount of inorganic carbon required for this amount of nitrification is quite low. McGhee(13) reported that the inorganic carbon requirements for the nitrite oxidation step could be met without inorganic carbon being present in the bulk solution. The utilizable source of inorganic carbon was the CO_2 generated from endogenous respiration within the biofilm.

The simplified oxidative reactions below describe the salient aspects of the microbial oxidation of ammonia. The microorganisms derive energy from these reactions; this energy is used for CO_2 fixation.

Ammonia Oxidation:

$$NH_4^+ + 1.5O_2 + 2HCO_3^- \longrightarrow NO_2^- + 2H_2CO_3 + H_2O \qquad (1)$$

Nitrite Oxidation:

$$NO_2^- + 0.5O_2 \longrightarrow NO_3^- \qquad (2)$$

Overall Reaction:

$$NH_4^+ + 2O_2 + 2\ HCO_3^- \longrightarrow NO_3^- + 2H_2CO_3 + H_2O \qquad (3)$$

As can be seen from Equation 1, the nitrification process results in the production of acid which neutralizes the alkalinity in the wastewater. Theoretically, 7.1 mg/l of alkalinity is destroyed for each 1 mg/l of ammonia oxidized. The destruction of alkalinity results in pH depression.

The actual pH depression is mitigated somewhat by the removal of carbonic acid through the stripping of CO_2 from the wastewater surface (10). However, under low alkalinity conditions, pH depression is enhanced due to the reduced buffer capacity of the wastewater. The level of alkalinity within wastewaters varies widely. The major factor influencing the amount of alkalinity present is the source of the carriage water, or the drinking water supply. High alkalinities normally are associated with ground water supplies and much lower alkalinities are associated with surface supplies. Domestic wastewater contributes from 50 to 200 mg $CaCO_3$/l to the natural alkalinity of the carriage water (14). Therefore, the amount of alkalinity in a domestic wastewater may range from less than 100 mg $CaCO_3$/l to several hundred mg $CaCO_3$/l. The net effect of such variations in alkalinity is to provide a different buffering capacity for each wastewater treatment system. Domestic wastewaters normally contain from 12 to 25 mg/l of ammonia-nitrogen. The range of alkalinity destroyed during the nitrification of these concentrations of ammonia is 85 mg $CaCO_3$/l to 178 mg $CaCO_3$/l. Obviously, the pH depression resulting from nitrification can be slight for low ammonia-high alkalinity wastewaters or significant for high ammonia-low alkalinity wastewaters. Wastewater pH levels are typically around pH 7.5. However, pH depression to below pH 7.0 is common for low alkalinity wastewaters.

The level of pH has an important effect on the nitrification process. There have been a number of researchers since the turn of the century who have addressed the subject of the effect of pH on nitrification. Those researchers who have made contributions pertinent to this study are listed in reverse chronological order in Table 1. It is interesting to note the large variation in the effect of pH on biological nitrification that is reported in the literature. The variation in the effect of pH on nitrification is due in large measure to the nature of the experiments undertaken, i.e., the homogeniety of culture involved (pure versus mixed culture); scale of the experiment (laboratory to full scale); nature of the biofilm (suspended versus fixed film) and a variety of (and frequently unspecified) acclimation times utilized in the experiments.

Several investigators (1, 3, 15, 16) within recent years have attempted to provide more information on nitrification within a fixed film mode. Haug and McCarty (15) utilized

Table 1. Literature Review of Optimum pH Values for Nitrification.

Ref.	Author	Year	Optimum pH	Organism or System Studied
3	Miller, et al.	1979	8.0-8.5	RBC (Pilot)
1	Borchardt, et al.	1978	7.1-8.6	RBC (Pilot)
13	McGhee	1975	8.0-9.0	A.S. (Lab)
17	Srna & Baggaley	1975	7.45	Sub. Filt.(Lab)
18	Hutton & LaRocca	1975	8.4-8.6	A.S.
16	Huang & Hopson	1974	8.4-9.0	Biofilm (Lab)
15	Haug & McCarty	1972	7.8-8.3	Sub. Filt.(Lab)
19	Mulbarger	1972	8.4	A.S.
20	Wild, et al.	1971	8.4	A.S.
21	Loveless & Painter	1968	7.5-8.0	Nitrosomonas
22	Downing & Knowles	1967	7.2-8.0	--
23,24	Andersen	1964	8.4-8.5	Nitrosomonas
25	Boon & Landelot	1962	7.0-8.6	Nitrobacter
26	Engel & Alexander	1958	7.0-9.0	Nitrosomonas
27	Bushwell & Shiota	1954	8.0-8.5	Nitrosomonas
28	Hoffman & Lees	1952	8.0-9.0	Nitrosomonas
29	Meyerhoff	1917	8.5-8.8	Nitrosomonas
			8.5-9.0	Nitrobacter

a laboratory scale fixed film submerged reactor and a synthetic wastewater and performed a short term pH-nitrification study (18 hours at each pH value) using a biofilm developed at neutral pH and observed essentially the same rate of nitrification at pH 6.5 as at pH 9.0. At pH 6.0, the observed rate of nitrification was reduced to approximately 42 percent of the maximum rate and nitrification essentially stopped at pH 5.5. However, after only 10 days of operation at pH 6.0, the submerged filter was reported to have acclimated sufficiently to perform at the maximum rate of nitrification. This finding demonstrates the ability of nitrifying organisms to acclimate to low pH condition. The reason for this unique finding may be due in part to non-equilibrium conditions existing within the submerged filter after the startup period. Huang and Hopson (16) utilized a laboratory scale inclined fixed film surface and a synthetic wastewater to evaluate the effect of pH on nitrification. Their experiment examined the short term (less than 10 hours) effect of pH on the nitrification process and produced a maximum rate of nitrification at pH 8.4 to pH 9.0 with approximately 25 percent of the maximum rate occurring at pH 6.0. After three weeks of acclimation at pH 6.6, the rate of nitrification was approximately 85 percent of the maximum rate observed.

Borchardt (1) performed a short term pH-nitrification study utilizing a 0.6 meter pilot RBC treating domestic wastewater effluent from a trickling filter in a laboratory where ammonia, alkalinity and pH were controlled. The rate of nitrification was examined at eleven different levels of alkalinity after a short but undefined acclimation period. The results of this short term study revealed a nearly constant rate of nitrification between pH 7.1 and 8.6. Approximately 25 percent of the maximum rate of nitrification was observed at pH 6.5 and zero nitrification was indicated at pH 6.0. Borchardt was careful to point out the limitations of attempting to extrapolate his short term data into the long term.

Miller (3) most recently reported on a pilot scale 0.5 meter RBC treating domestic wastewater effluent from a pilot trickling filter wherein significantly greater rates of nitrification were observed at elevated pH levels (pH 8.0 to pH 8.5) than at neutral pH (approximately pH 7.1). This nitrification study is unique in that lime addition for phosphorus removal preceded the nitrification process and the nitrifying RBC stages had acclimated at

the elevated pH levels. A transition in biofilm performance was observed when the elevated pH of the wastewater was reduced to the neutral pH range. Nitrification performance initially remained unchanged. After approximately four days, the performance level started to deteriorate. In nine days, the performance had reverted to a lower nitrification level. This latter finding was not discussed fully by Miller; however, it is important because it helps to establish potential physical differences between the biofilms developed at neutral and elevated pH levels. This situation was not observed by the other investigators using fixed films mentioned above because none ever attempted to acclimate biofilms at the elevated pH levels. Such differences cannot be assumed to be purely indicative of only the pH dependent rates of microbial nitrification. These differences also are reflective of the entire heterogenous population developed within each biofilm which dictate film development, cohesion, and retention characteristics (sludge age). There is essentially no information within these wastewater nitrification studies which addresses changes in biofilm and microbial populations under various pH conditions. In general, this important consideration has been ignored in such wastewater research studies. However, current research efforts such as those by Olem (30), LaMotta (31) and Characklis (32) are starting to examine more closely the mechanics of biofilm development and the characterization of microbial populations.

The addition of alkaline chemicals to wastewater treatment systems to increase pH and provide added buffer capacity has been attempted with varying degrees of success. Heidman (33) conducted a pilot study at the Blue Plains WWTP using an activated sludge system which incorporated pH controlled nitrification. This study was inconclusive because it failed to demonstrate the relative nitrification without chemical addition. Hutton (18) demonstrated the feasibility of optimizing the nitrification of high ammonia strength industrial wastewaters with alkaline chemical addition. Lue-Ling (34) reported success in using alkaline chemical addition to nitrify high ammonia strength lagoon supernatant with RBCs. Hitdlebaugh (35) attempted to enhance the nitrification of domestic wastewater with RBCs through alkaline chemical addition; however, the results were inconclusive. The literature fails to address the efficacy of optimizing domestic wastewater

nitrification within the RBC system through pH control as well as the use of alternative pH control schemes.

OBJECTIVES AND SCOPE

The objectives of this research were to:

1. Establish the relative rates of nitrification for domestic wastewater treatment within an acclimated RBC fixed film system as a function of pH.
2. Observe and characterize the relative changes in the RBC biofilm as a function of pH.
3. Evaluate the efficacy of chemical addition to improve nitrification within an RBC fixed film system through the maintenance of an optimum pH.
4. Evaluate alternative alkaline chemicals for pH controlled nitrification for the RBC.
5. Develop design criteria, as appropriate, for pH controlled nitrification for the RBC.

EXPERIMENTAL PROCEDURES

Pilot scale 0.5 meter diameter RBC systems were used in this research to nitrify high rate trickling filter effluent from the Pennsylvania State University (PSU) wastewater treatment plant (WWTP). The effect of pH on nitrification within an RBC was evaluated using four single stage RBCs operating in parallel (Figure 1). The pH of the RBC systems treating the PSU WWTP trickling filter effluent was varied first from pH 6.3 to pH 7.5 and then from pH 7.6 to pH 8.8 and the relative levels of nitrification at the various pH's then were observed. High pH and low pH environments were created by adding sodium hydroxide and sulfuric acid, respectively, to the wastewater streams after clarification (Figure 1). Each observation period was started with no biofilm on the RBC discs and lasted approximately ten weeks.

The effect of alkaline chemical addition was evaluated utilizing five 2-stage RBCs operating in parallel (Figure 2). The level of nitrification of a low pH 2-stage control RBC system (control) was compared against the nitrification level of four other 2-stage RBC systems receiving four different alkaline chemicals. The four alkaline chemicals used were calcium hydroxide, sodium carbonate, sodium hydroxide, and sodium bicarbonate.

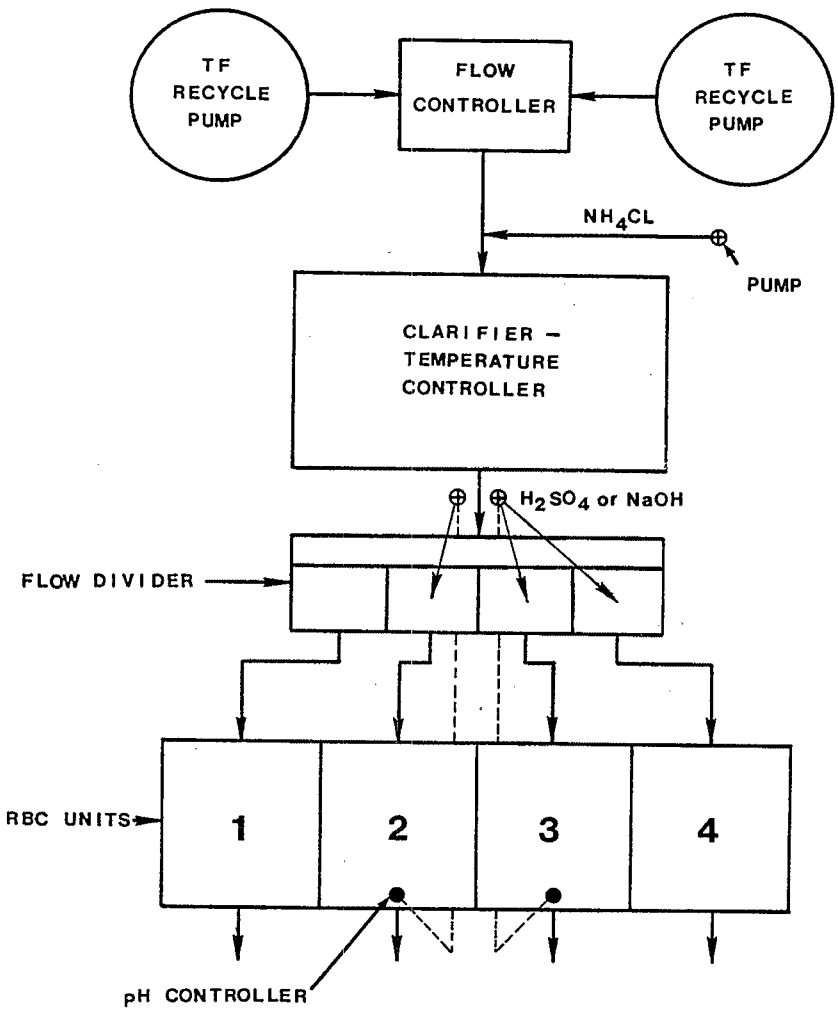


Figure 1. Schematic Diagram of the Pilot RBC Units for the Low pH- and High pH-Nitrification Study

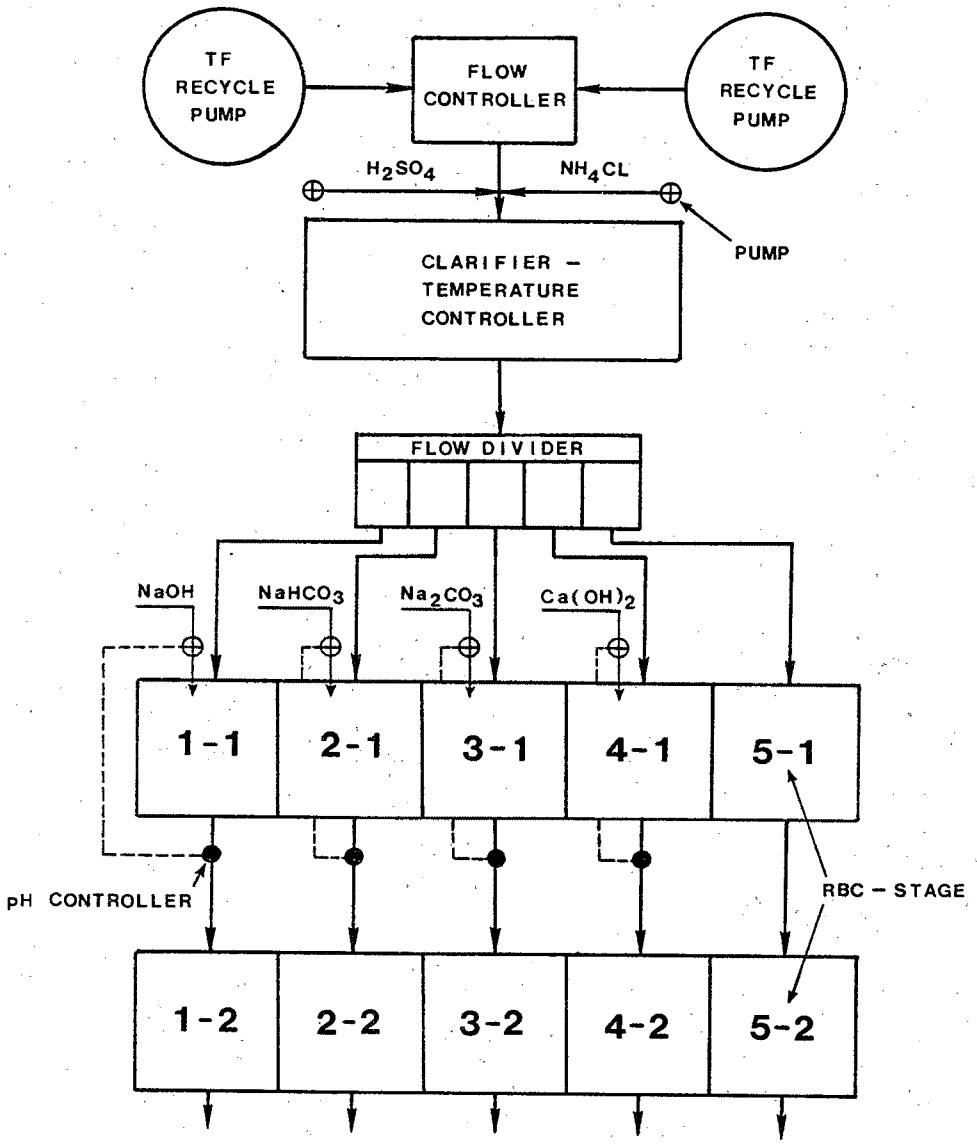


Figure 2. Schematic Diagram of the 2-Stage RBC Systems of the Alkaline Chemical Addition Study.

The alkaline chemicals were added to the first stage only. The first stages of the calcium hydroxide, sodium carbonate, and sodium hydroxide RBC systems were maintained at the optimum pH level for nitrification (approximately pH 8.5). The first stages of the sodium bicarbonate and the control RBC systems were maintained at pH 7.5 and pH 7.0, respectively. The low pH wastewater was created by adding sulfuric acid prior to clarification and the alkaline chemicals then were added directly to the first stage of each RBC. Ammonium chloride was added to the wastewater during PSU break periods to augment the low influent ammonia-nitrogen. There was no biomass on the discs at the start of the test. The observation period for this part of the study lasted approximately 11 weeks.

Wastewater sampling was accomplished by compositing grab samples on influents and effluents. Biofilm sampling and analyses were performed in accordance with modifications of the procedures reported by Olem (30). The most probable numbers (MPN) of ammonia-oxidizing bacteria were determined using a modification of the Nitrosomonas MPN technique of Alexander and Clark (36) which was reported by LaBeda and Alexander (37) as well as Rowe (38). The nitrite-oxidizing bacteria MPN values were determined using a modification of the Nitrobacter MPN technique of Alexander and Clark (36) which was reported by LaBeda and Alexander (37) and Ghiorse and Alexander (39). The enumeration of heterotrophic bacteria was accomplished by spread plating serial dilutions of Modified Taylor's Media (40). Detailed descriptions of all sampling and analytical procedures are found in Stratta (41).

RESULTS AND DISCUSSION

Low pH-Nitrification Study

This research phase was devoted to the evaluation of the relative rates of nitrification in single stage RBC systems operated at pH 7.5, 7.1, 6.5 and 6.3. The pH 7.5 RBC treated the unaltered wastewater and served as the control. Data showing the operational characteristics of the four RBC systems are presented in Table 2. The rates of nitrification which initially developed for the pH 7.5 and the pH 7.1 RBC systems were greater than the nitrification rates of the two lower pH RBC systems. Nitrification was not established in the pH 6.3 RBC system

Table 2. Pilot Single-Stage Nitrifying RBC Operating Characteristics

Secondary Clarifier	
Surface Settling Rate (@6.8 m^3/d)	5.9 $m^3/d \cdot m^2$
Detention Time (@6.8 m^3/d)	2.1 hr
RBC	
Number of RBCs	4
Stages per RBC	1
Discs per Stage	9
Disc Diameter	0.5 m
Disc Area - Total	5.3 m^2
Rotational Speed	13 rpm
Peripheral Speed	0.34 m/sec
Hydraulic Loading[a]	81 $l/m^2 \cdot d$

[a]The hydraulic loading for all four RBC units was nominally 81 $l/m^2 \cdot d$ (2 $gal/d \cdot ft^2$). The hydraulic loading calculation is based upon the assumption that each RBC is the first stage of a 4-stage RBC.

during the first 25 days of operation. Two short-term pH excursions may have had an adverse impact on nitrification development on the pH 6.3 RBC. On Day 27, the pH of the pH 6.3 unit was adjusted upward to pH 6.7 in an attempt to obtain additional information regarding the nitrification rate between pH 6.5 and pH 7.1; this RBC is referred to hereafter as the 6.3/6.7 RBC. Based upon nitrification performance, a period of relative equilibrium was established by about Day 37 for the pH 7.5, 7.1, and 6.5 RBC units. Data on the relative amounts of ammonia-nitrogen removed by the RBCs are presented in Table 3. The rate of nitrification of the pH 7.1 RBC system was 96 percent of that observed at pH 7.5 and the rate for the pH 6.5 RBC was 80 percent of the rate for the pH 7.5 RBC. Because of the long period of time required for nitrification to become established in the pH 6.3/6.7 RBC, data for this unit are not included in Table 3. Nitrogen balances for each

Table 3. Relative Rates of Nitrification for RBC Systems Operating Under Low pH Conditions[a]

RBC pH	Ammonia-N Removed g NH_3-N/m^2·d	Percent of Maximum
7.5	2.5	100
7.1	2.4	96
6.5	2.0	80

[a]Based upon data from Day 37 to Day 69

of the RBC systems are presented in Table 4. The slightly lower nitrogen recoveries for the two higher pH systems might reflect small nitrogen losses associated with denitrification within the heavier biofilms as well as minor losses due to ammonia stripping.

Table 4. RBC Nitrogen Balances for the Low pH-Nitrification Study[a]

RBC	pH	Total Nitrogen[b] - mg/l Influent	Effluent	Percent Recovery
1	7.5	19.5(23)[c]	19.3(23)	99
2	7.1	19.6(23)	19.7(23)	100
3	6.5	19.7(24)	20.3(24)	103
4	6.3/6.7	19.6(23)	20.2(23)	103

[a]Based upon data from Day 37 to Day 69

[b]Total nitrogen is total oxidized nitrogen plus total Kjeldahl nitrogen (TKN)

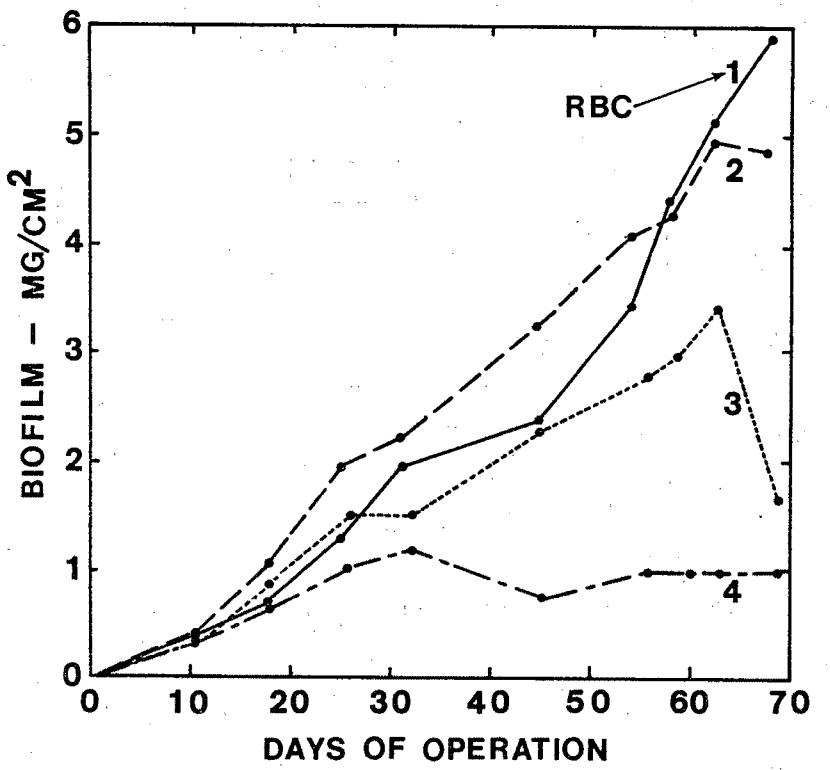
[c]Number in parenthesis is the number of samples utilized in the total nitrogen determinations.

The four RBCs developed biofilms which could be sensed by touch within 48 hours. A noticeable bronze color developed after five days of operation. By the tenth day, all RBC systems had developed thin and highly uniformly textured coatings which possessed a visually apparent gradation. The heaviest biofilm appeared in the pH 7.5 RBC and the lightest growth of biofilm was in the pH 6.3/6.7 RBC. All four RBC units showed some degree of sloughing by Day 19 with the greatest sloughing occurring in the pH 6.3/6.7 RBC. The biofilm color changed from bronze to brown with increasing age and increased biomass. After the loss of the initial biofilm uniformity, the biofilm became increasingly patchy with time and decreasing pH. At the conclusion of this phase of the study, the non-uniformity of the biofilm was quite evident visually and seemed to be related directly to the relative nitrification rates recorded over the duration of the study. The

patchy appearance was attributed mainly to biofilm loss resulting from hydraulic shear. However, biofilm sloughing from the disc surface did occur.

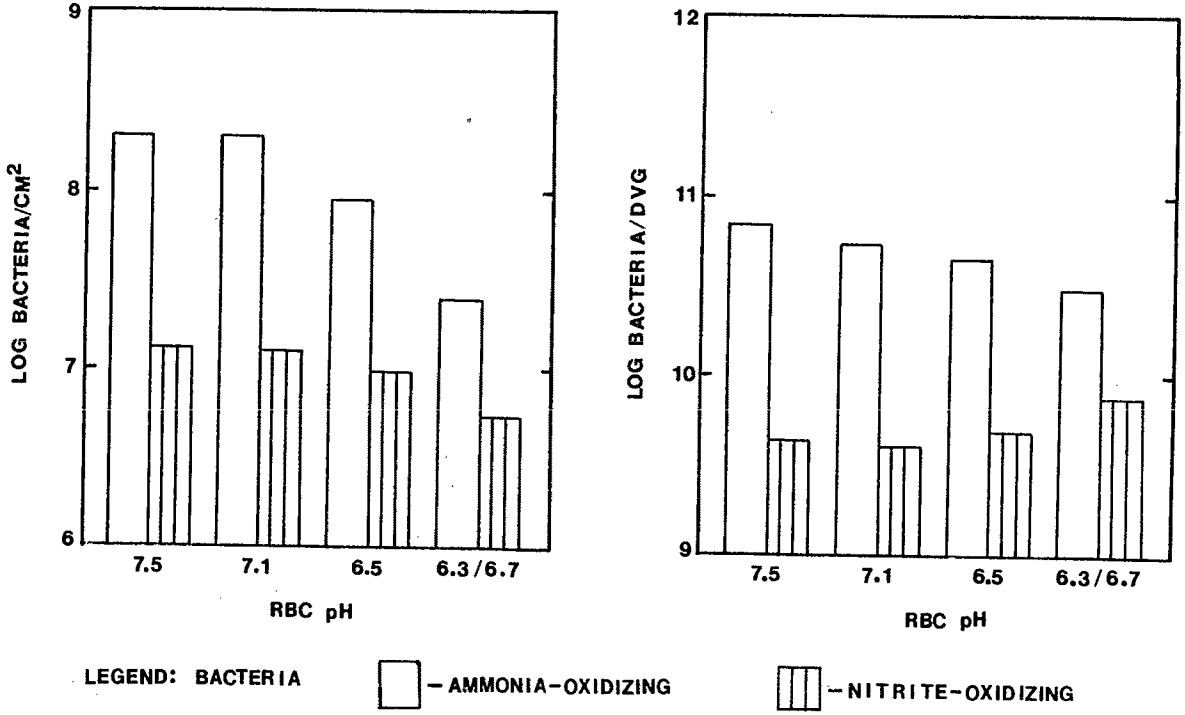
The RBC disc biofilm development data for all four RBC systems during the low pH-nitrification study are presented in Figure 3. The RBC systems at pH 7.5 and 7.1 showed the best performance and had the most disc biofilm. The pH 6.5 RBC showed a lower level of performance and less biofilm. The RBC which was operated initially at pH 6.3 and later adjusted to pH 6.7 had the lowest performance level throughout most of the 69-day study and also developed the least amount of biofilm. The maximum ammonia-oxidation levels for the pH 7.5, 7.1, and 6.5 RBC's were achieved when the biofilm masses were approximately 2.0, 2.2, and 1.5 mg/cm^2 respectively. Increases in disc biomass did not enhance the nitrification rates for any of these RBC systems. The pH 6.3/6.7 RBC added biofilm during the first three weeks of operation at a rate comparable to that of the pH 7.5 RBC yet showed no nitrification capacity. This result indicates that, at least initially, organisms other than nitrifying bacteria were inhabiting the RBC discs.

The relative geometric mean data for the nitrifying bacterial populations per unit disc area and per unit volatile weight for each RBC system after the initial month of startup are presented in Figure 4. These graphs demonstrate clearly that the total number of viable nitrifying bacteria on each RBC was related directly to overall RBC nitrification performance. The higher pH systems had larger populations of both ammonia-oxidizing and nitrite-oxidizing bacteria. The sustained depressed nitrification performance and the relatively low nitrifying bacteria populations of the 6.3/6.7 RBC indicated that a significant period of time was required for complete autotrophic adjustment in response to system changes under relatively low pH conditions. The ratios of ammonia-oxidizing bacteria to nitrite-oxidizing bacteria for the pH 7.5, 7.1, 6.5 and 6.3/6.7 RBC units were 16:1, 14:1, 9.4:1 and 3.3:1, respectively. This observation indicates that the lower pH systems favor nitrite-oxidizing bacteria relative to ammonia-oxidizing bacteria. This conclusion is shown graphically in Figure 4 which shows that the number of ammonia-oxidizing bacteria per dvg (dry volatile gram) decreased with decreasing pH, while the number of nitrite-oxidizing bacteria per dvg increased for the two



LEGEND: RBC ——— 1-pH 7.5 – – – 2-pH 7.1
····· 3-pH 6.5 —·— 4-pH 6.3/6.7

Figure 3. RBC Disc Biofilm Development for the Low pH-Nitrification Study



NOTE: The MPN data are geometric means of weekly enumerations starting after Day 30 and ending on Day 69.

Figure 4. Relative Geometric Mean Populations of Ammonia-Oxidizing and Nitrite-Oxidizing Bacteria on a Unit Area and Unit Dry Volatile Weight Basis for the Low pH-Nitrification Study

lower pH RBC units. The volatile contents for these biofilms were 82, 83, 84, and 81 percent for the pH 7.5, 7.1, 6.5, and 6.3/6.7 RBCs, respectively.

High pH-Nitrification Study

This research phase was devoted to the evaluation of the relative rates of nitrification in single stage RBC systems operated at pH 7.6, 8.0, 8.5, and 8.8. The pH 7.6 RBC treated the unaltered wastewater and served as the control. The operational characteristics of the four RBC systems were the same as reported previously in Table 2. On Day 36, a pH excursion to approximately pH 11.0 for an estimated two hours occurred within the pH 8.8 RBC with rather dramatic results. This short-term transient condition appeared to have little effect on the ammonia-oxidation process. However, the RBC experienced an immediate loss of nitrite-oxidation capability and a very slow nitrite-oxidation recovery. Based upon ammonia removal, and not complete oxidation, a period of relative equilibrium was established after approximately five weeks of operation. Data on the relative amounts of ammonia-nitrogen removed by the RBCs are presented in Table 5. The pH 8.8 RBC

Table 5. Relative Rates of Nitrification for RBC Systems Operating Under High pH Conditions[a]

RBC pH	Ammonia-N Removed ($g\ NH_3\text{-}N/m^2 \cdot d$)	Percent of Maximum
7.6	2.0	65
8.0	2.6	84
8.5	3.1	100
8.8	2.9	94

[a]Based upon data from Day 38 to Day 71

and the pH 8.0 RBC removed 94 and 84 percent as much ammonia as did the pH 8.5 system, respectively; whereas the control RBC removed only 65 percent as much ammonia. Data on the nitrogen balances for the four RBC systems for this time period are presented in Table 6. As noted earlier, the slightly lower nitrogen recoveries at the higher pH conditions may be due to small nitrogen losses resulting from ammonia stripping and denitrification within the heavier biofilms. These nitrogen recovery results indicate that ammonia stripping was not a major factor affecting the change in ammonia levels between pH 7.6 and pH 8.8.

Table 6. RBC Nitrogen Balances for the High pH Nitrification Study[a]

RBC	pH	Total Nitrogen[b] - mg/l Influent	Effluent	Percent Recovery
1	7.6	21.7(21)[c]	21.1(21)	97
2	8.0	21.5(21)	20.2(21)	94
3	8.5	21.7(21)	20.6(21)	95
4	8.8	21.8(21)	20.8(21)	95

[a]Nitrogen balances are based upon data from Day 38 to Day 71.

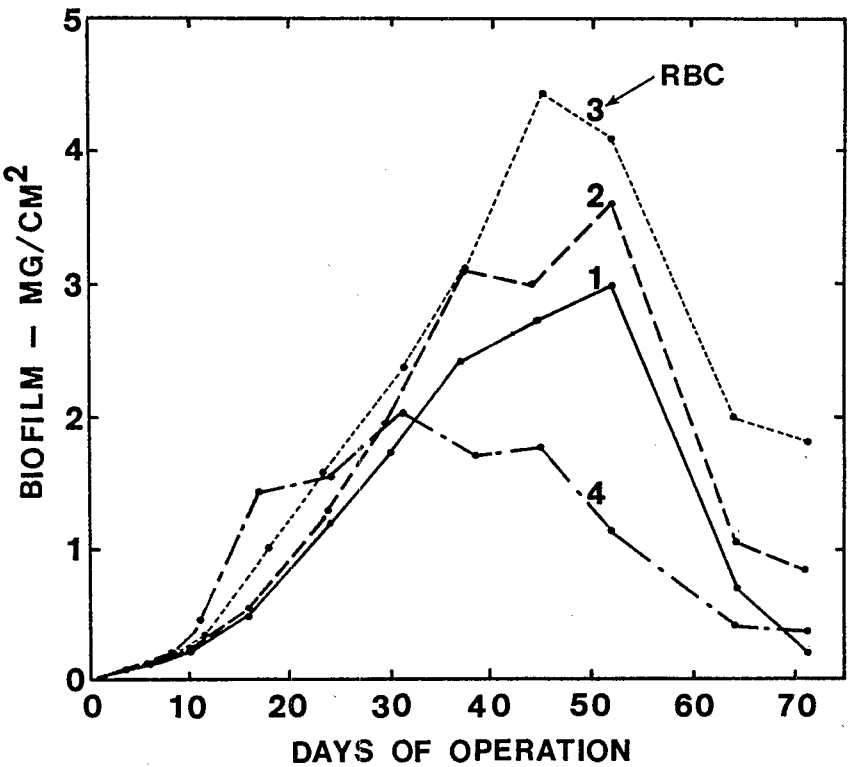
[b]Total nitrogen is total oxidized nitrogen plus total Kjeldahl nitrogen (TKN).

[c]Number in parenthesis is the number of samples utilized in the total nitrogen evaluations.

The four RBCs developed biofilms which could be sensed by touch within 48 hours. A noticeable reddish-brown biofilm was evident on all the discs by the third day of

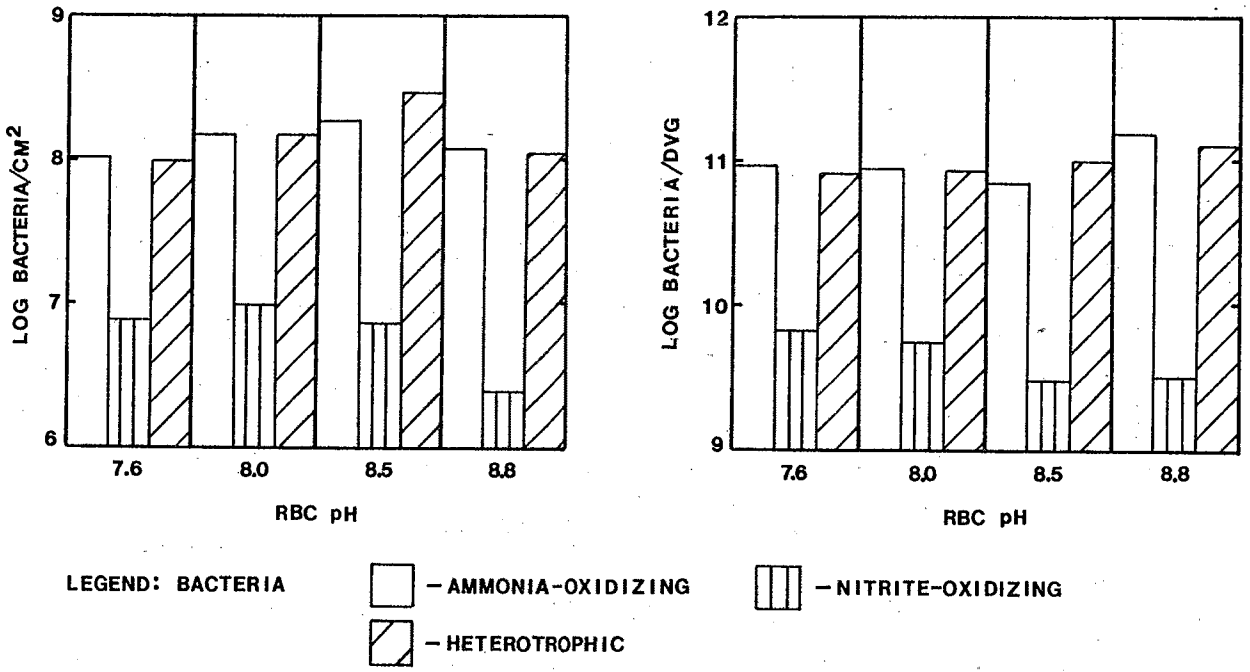
operation. By the eighth day, the four RBCs had developed thin and highly uniformly textured biofilms which possessed a visually apparent gradation. The pH 8.5 and pH 8.8 RBC biofilms initially developed more rapidly than did the pH 7.6 and pH 8.0 RBC biofilms. This initial biofilm gradation was not related to ammonia removal efficiency. The biofilm color had changed from reddish-brown to tan or bronze on all discs by the 10th day. By the 13th day of operation, the two lower pH RBCs had the most uniformly textured biofilms while the pH 8.5 and pH 8.8 RBCs were developing a "dimpled" appearance associated with the heavier biofilms. The pH 8.8 RBC developed a patchy appearance and also had started to slough significantly after only two weeks. As time progressed, the RBC systems added biofilm, but their texture became less uniform. The heavier biofilms appeared to be associated with the pH 8.0 and pH 8.5 systems. The pH 8.8 RBC experienced the greatest biofilm sloughing. Figure 5 presents the RBC disc biofilm development data for all four RBC systems during the high pH-nitrification study. After the initial month of operation, the levels of disc biofilm for the pH 7.6, 8.0, and 8.5 RBCs were related directly to their relative level of performance. The pH 8.8 RBC experienced an initially high rate of biofilm development; however, it reached a peak mass per unit area concentration on Day 31 and then experienced a continuous biofilm loss. All four RBC systems experienced a marked decline in disc biofilm after the end of the PSU spring term on Day 52 when the influent CBOD concentration decreased. The four RBC systems appeared to achieve an initial maximum level of nitrification performance in about three weeks. These performance levels cooresponded to disc biofilm concentrations of approximately 0.8, 1.0, 1.2 and 1.4 mg/cm^2 for the pH 7.6, 8.0, 8.5, and 8.8 RBCs, respectively. As demonstrated previously, the increase in disc biomass did not improve the rate of nitrification for any of the systems.

The data on the relative geometric mean bacteria populations per unit of disc area and per unit weight of dry volatile biofilm for each RBC system after the initial 30 days of operation are presented graphically in Figure 6. The total numbers of ammonia-oxidizing and heterotrophic bacteria increased with increasing pH up to pH 8.5 and then experienced a drop at pH 8.8. The total number of nitrite-oxidizing bacteria was greatest at pH 8.0 and decreased at pH 8.5 and pH 8.8. The number of ammonia-oxidizing bacteria relative to the total biofilm population was similar for



LEGEND: RBC ——1-pH 7.6 – – – –2-pH 8.0
·····3-pH 8.5 —·—4-pH 8.8

Figure 5. RBC Disc Biofilm Development for the High pH-Nitrification Study



NOTE: Geometric means are based upon six sets of enumerations for each bacteria classifications from Day 30 to Day 71. Nitrifying enumerations are MPN values and heterotrophic enumerations are plate counts.

Figure 6. Relative Geometric Mean Populations of Ammonia-Oxidizing, Nitrite-Oxidizing, and Heterotrophic Bacteria on a Unit Area and Unit Weight of Dry Volatile Disc Biofilm Basis for the RBCs of the High pH-Nitrification Study

the pH 7.6, 8.0, and 8.5 systems but greatest at pH 8.8. Similarly, the heterotrophic population increased with pH. However, the nitrite-oxidizing bacteria populations were nearly identical at pH 7.6 and pH 8.0 but lower at pH 8.5 and pH 8.8. The ratios of heterotrophic to ammonia-oxidizing to nitrite-oxidizing bacteria based on the data presented in Figure 6 for the pH 7.6, 8.0, 8.5, and 8.8 RBC units were 12:15:1, 16:16:1, 37:24:1, and 43:48:1, respectively. In general, these population figures indicate that the heterotrophic bacteria and the ammonia-oxidizing bacteria are favored over the nitrite-oxidizing bacteria with respect to increasing pH. During this period, the mean biofilm concentrations were 1.80, 2.26, 2.99, and 1.23 mg/cm^2 for the pH 7.6, 8.0, 8.5, and 8.8 RBC units, respectively. The volatile content was 86, 86, 83, and 75 percent for the pH 7.6, 8.0, 8.5, and 8.8 RBC units, respectively. This lower volatile content of the pH 8.8 RBC was attributed to low level precipitation of calcium carbonate and entrainment of the precipitate within the disc biofilm.

At the conclusion of the 10-week high pH-nitrification study, the two RBC systems which had been operating at pH 7.6 and pH 8.5 were utilized in a short-term pH-nitrification study wherein the two RBC systems experienced simultaneous short term changes in pH, i.e. 2 hours of operation at each pH level. The pH level started at pH 9.0 and was decreased progressively downward to pH 6.0 without interruption. Alkalinity and pH levels were maintained in each RBC by direct feed of sodium hydroxide and sulfuric acid solutions. This test was run twice (Day 73 and Day 79) with similar results. The average values of nitrogen removal data obtained from these two runs are shown in Figure 7.

The shapes of the performance curves for the two RBCs are similar yet the level of nitrification for the two RBCs are markedly different. Clearly, the RBC which had acclimated at pH 8.5, and had a history of elevated performance, retained its higher performance level in the short-term and continued to perform significantly better than the biofilm acclimated at pH 7.6. The RBC response to short-term changes in pH is relatively constant between pH 7.9 and pH 9.0 but highly dependent upon the previous acclimated level of nitrification for the given RBC biofilm. Data on the alkalinity levels also have been included in Figure 7. The amount of alkalinity present at the low

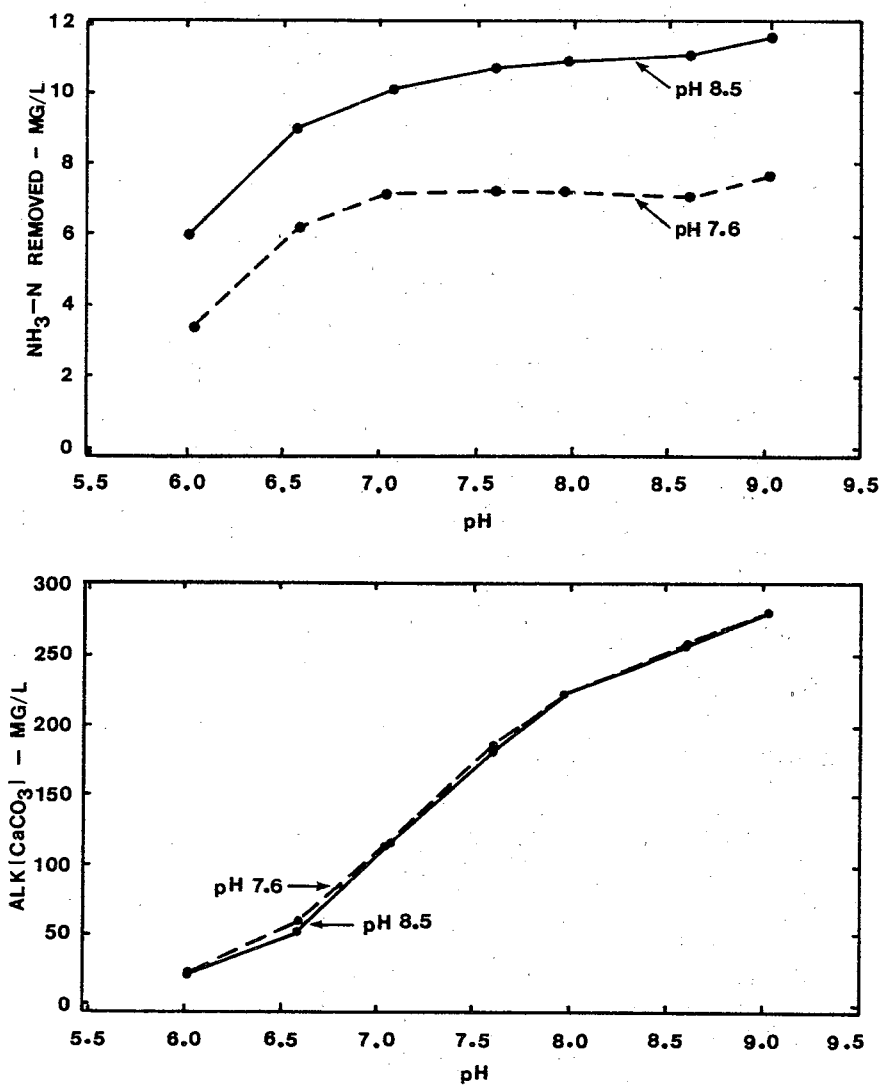


Figure 7. The Relative Rates of Nitrification of RBC Systems Acclimated at pH 8.5 and pH 7.6 and Subjected to Short Term pH Changes and Related Alkalinity Levels

pH levels demonstrates that even at 20 mg/1 of $CaCO_3$, significant amounts of nitrification were achieved. The response of the lower pH system closely resembles the data of Borchardt (1) and similarly reveals good nitrification at very low alkalinity levels. This result tends to reinforce the observation that the pH level is much more important than the alkalinity level per se in terms of effect on nitrification. The amount of time required for an RBC system to adjust to an altered pH condition is discussed further in Stratta (41).

Alkaline Chemical Addition Study

This phase of the research was devoted to the evaluation of the rates of nitrification of 2-stage RBC systems maintained at elevated pH levels through alkaline chemical addition. The pH and alkalinity levels within the first stages of four RBC systems were adjusted upward and maintained artificially with four different alkaline chemicals. Calcium hydroxide, sodium carbonate, and sodium hydroxide were used to maintain approximately pH 8.5, the optimum pH, in the first stage of three different RBCs. Sodium bicarbonate was used to maintain pH 7.5 in the first stage of the fourth RBC. A fifth RBC which treated low pH-low alkalinity wastewater was used as a control. Data on the operational characteristics of the five RBC systems are presented in Table 7.

The overall nitrification capacity of the 2-stage RBC control system developed more slowly than did that of the higher pH systems. However, the control RBC and the alkaline chemical feed RBC systems were operating at approximately the same level of nitrification performance after a little more than three weeks of operation. The levels of performance for all five systems were very similar for about the next ten days. The overall performance of the control system, operating at the lower pH level, started to deteriorate after approximately 35 days of operation.

Based upon the overall nitrification performance of the five 2-stage RBC systems, a period of relative equilibrium was established by about Day 38. The data on the relative amounts of total ammonia-nitrogen removed by the five RBCs from Day 38 to Day 75 are presented in Table 8. These data show that the overall ammonia-nitrogen removals for the sodium hydroxide, sodium carbonate, and

Table 7. Pilot 2-Stage Nitrifying RBC Systems' Operating Characteristics

Secondary Clarifier		
Surface Settling Rate (@ 8.6 m^3/d)	-	7.4 $m^3/d \cdot m^2$
Detention Time (@ 8.6 m^3/d)	-	1.7 hr
RBC		
Number of RBCs	-	5
Stages per RBC	-	2
Discs per Stage	-	9
Disc Diameter	-	0.5 m
Disc Area - Total	-	10.6 m^2
Rotational Speed	-	13 rpm
Peripheral Speed	-	0.34 m/sec
Hydraulic Loading[a]	-	81 $l/m^2 \cdot d$

[a]The hydraulic loading for all five RBC units was nominally 81 $l/m^2 \cdot d$ (2 $gal/d \cdot ft^2$). The hydraulic loading calculation is based upon the assumption that each 2-stage RBC system contained the first two stages of a 4-stage RBC.

Table 8. Relative Rates of Nitrification for the RBC Systems of the Alkaline Chemical Addition Study[a]

RBC	Alkaline Chemical	Ammonia-N Removed (g NH_3-N/m^2·d)	Percent Removed	Percent of Maximum
1	NaOH	2.52	86	99
2	$NaHCO_3$	2.40	82	94
3	Na_2CO_3	2.54	87	100
4	$Ca(OH)_2$	2.55	87	100
5	Control	2.14	73	84

[a]Based upon data from Day 38 to Day 75

calcium hydroxide RBC systems were nearly the same. The performance level of the sodium bicarbonate RBC system was about 6 percent less than that of the other three high pH systems. The control RBC, which was operated at the lowest pH conditions, removed about 16 percent less ammonia-nitrogen than did the three high pH alkaline chemical feed systems. Table 9 presents the ammonia removal data for each respective RBC stage. The data on the amounts of ammonia-nitrogen removed clearly demonstrate that the greatest removal occurs at the elevated pH conditions. The amount of ammonia removed by the first and second stages of the three high pH alkaline chemical feed RBC systems was essentially the same. The sodium bicarbonate RBC system had lower pH levels and lower performance in both stages. The control had the lowest stage pH levels and the poorest performance for both stages. Nitrogen balances during this period for the stages of the five RBC systems are presented in Table 10. These nitrogen balances follow the same pattern previously reported. The percent recovery decreased slightly as pH increased. Again, this lower

Table 9. Relative Rates of Nitrification for the Stages of the RBC Systems of the Alkaline Chemical Addition Study[a]

RBC-Stage	Alkaline Chemical	pH	Ammonia-N Removed (g NH_3-N/$m^2 \cdot d$)	Percent Removed	Percent of Maximum
1-1	NaOH	8.5	2.53	43	98
2-1	$NaHCO_3$	7.5	2.33	40	91
3-1	Na_2CO_3	8.4	2.55	44	99
4-1	$Ca(OH)_2$	8.5	2.57	44	100
5-1	Control	7.0	2.14	37	83
1-2	-	7.9	2.50	77	98
2-2	-	7.7	2.46	72	97
3-2	-	8.0	2.52	78	99
4-2	-	7.9	2.54	78	100
5-2	-	6.9	2.14	59	84

[a]Based upon data from Day 38 to Day 75

[b]Based upon ammonia-nitrogen influent to each RBC stage.

[c]Based upon maximum ammonia-nitrogen removed by calcium hydroxide RBC stages.

Table 10. RBC Nitrogen Balances for the Alkaline Chemical Addition Study[a]

RBC-Stage	Alkaline Chemical	pH	Total Nitrogen[b] mg/l	Percent Recovery	
				Stage[d]	RBC
INFLUENT	-	6.5	24.8(24)[c]	-	-
1-1	NaOH	8.5	24.4(26)	98	-
2-1	$NaHCO_3$	7.5	24.0(24)	97	-
3-1	Na_2CO_3	8.4	23.5(25)	95	-
4-1	$Ca(OH)_2$	8.5	23.7(25)	96	-
5-1	Control	7.0	24.4(25)	98	-
1-2	-	7.9	22.8(24)	93	92
2-2	-	7.7	23.0(25)	96	93
3-2	-	8.0	22.3(25)	95	90
4-2	-	7.9	22.3(25)	94	90
5-2	-	6.9	23.8(25)	98	96

[a] Based upon data from Day 38 to Day 75.

[b] Total nitrogen is total oxidized nitrogen plus total Kjeldahl nitrogen (TKN).

[c] Number in parenthesis is the number of samples utilized in the total nitrogen determinations.

[d] Nitrogen balances for the stages are based upon stage influent nitrogen.

recovery is attributed to small ammonia losses due to ammonia stripping and denitrification within the thicker biofilms associated with the higher pH levels.

Biofilm which could be sensed by touch had developed on the discs of all stages within 36 hours. Within 72 hours from startup, all first stage discs had developed biofilms which were noticeably heavier than the second stage biofilms. The characteristic tan color associated with nitrifying biofilms had developed by Day 4 and was more apparent in the first stage biofilms. All of the biofilms were very uniform in texture. This initial biofilm growth appeared to be heavier than the biofilms developed during the previous research phases. By Day 6, the trough walls also had developed noticeable amounts of biofilm. Although both stage biofilms got heavier and darker with time, the heavier and darker biofilms were on the first stage discs. The first stage biofilms became brown while those on the second stage discs remained tan to bronze in color. By Day 16, discs in all the first stages had experienced some sloughing while those in the second stages retained their uniformity. The loss of uniformity in the second stages commenced about Day 21. As time progressed, the loss of biofilm uniformity was greatest for the control and the sodium bicarbonate RBC systems. The first stages of the calcium hydroxide, sodium carbonate, and the sodium hydroxide RBC systems had the heaviest and most uniform biofilm coatings. The biofilm uniformity related directly to the ammonia removal performance levels of the RBC systems. The loss of biofilm during the 75-day period was attributed mainly to hydraulic shear starting on the surface of the biofilm and progressing inward. Biofilm sloughing from the bare disc outward did not occur continuously. The former method of sloughing appeared to be associated with relatively low and steady CBOD loadings, while the latter form of sloughing appeared to be associated with relatively high and fluctuating CBOD loadings.

The data for the RBC biofilm concentrations, percent volatile matter, and percent nitrogen are presented in Table 11. These data show that the highest biofilm concentrations were associated with the higher pH levels. Only the addition of calcium hydroxide resulted in an increase in inert matter entrained within both the first and second stage biofilms and a significant increase in effluent suspended solids. Sodium hydroxide, sodium carbonate, and sodium bicarbonate additions did not affect the biofilm

Table 11. Mean Biofilm Concentrations, Percent Volatile Matter, and Percent Nitrogen in the RBC Disc Biofilm of the Alkaline Chemical Addition Study[a]

RBC-Stage	Alkaline Chemical	Stage pH	Biofilm mg/cm^2	Volatile Biofilm %	Nitrogen[b] %
1-1	NaOH	8.5	2.45	86	5.7
2-1	$NaHCO_3$	7.5	2.08	86	5.7
3-1	Na_2CO_3	8.4	2.51	87	7.5
4-1	$Ca(OH)_2$	8.5	3.03	66	7.0
5-1	Control	7.0	1.57	87	7.6
1-2	--	7.9	0.97	89	5.4
2-2	--	7.7	0.84	90	7.6
3-2	--	8.0	0.96	89	6.4
4-2	--	7.9	1.88	70	6.0
5-2	--	6.9	0.87	90	6.8

[a]Samples taken at weekly intervals from Day 36 to Day 75

[b]Nitrogen percentages are based upon weekly samples from Day 23 to Day 75.

volatile content; however, a slight increase in volatile content in all the second stage biofilms was noted. The addition of sodium hydroxide, sodium bicarbonate, and sodium carbonate caused only a slight increase of from 1 to 3 mg/l in the suspended solids in the RBC effluents; whereas the use of calcium hydroxide increased the effluent suspended

solids by approximately 20 mg/1. The observed increase in the calcium hydroxide RBC biofilm inert content as well as the increase in suspended solids is attributed to the reaction between the calcium hydroxide and the carbonic acid or carbon dioxide in the wastewater to form calcium carbonate.

The populations of ammonia-oxidizing, nitrite-oxidizing, and heterotrophic bacteria were monitored for both stages of each RBC system. Figures 8 and 9 present graphically the data on the relative geometric mean bacteria populations per unit of disc area and per unit weight of dry volatile biofilm for the stages of each RBC system from Day 36 to Day 74. This time period corresponds to the same period over which the relative nitrification rates are compared in Tables 8 and 9. The populations of all three groups of bacteria per unit area were greater for the first stages of the three high pH, high performance systems than for the first stages of sodium bicarbonate and control RBC systems. The first stages of the former were maintained at pH 8.4 to pH 8.5 while the latter were maintained at pH 7.5 and 7.0 for the sodium bicarbonate and control RBC systems, respectively. In the second stages, where there was less CBOD, less disc biofilm, and no pH control, the population differences were not as dramatic. The ratios of the populations for the three groups of bacteria for both stages of each RBC system are presented in Table 12.

Results of this research effort had indicated throughout the various phases that heterotrophic activity and biofilm development were enhanced under elevated pH conditions. In order to provide additional information regarding this observation, approximately 400 cm^2 of new disc material was added to the first stage discs of the control (pH 7.0), the sodium bicarbonate (pH 7.5), and the sodium hydroxide (pH 8.5) RBC systems on Day 62. The development of biofilm and the establishment of heterotrophic populations on these discs were monitored through Day 77. The resulting data are presented in Figures 10 and 11. The data demonstrated that both the biofilm and the heterotrophic activity developed more rapidly as pH increased from pH 7.0 to pH 8.5. During this 15-day test period, the influent CBOD (soluble and inhibited) concentration was approximately 8 mg/1. However, significantly greater amounts of CBOD, if present, may overshadow the more subtle influence of pH.

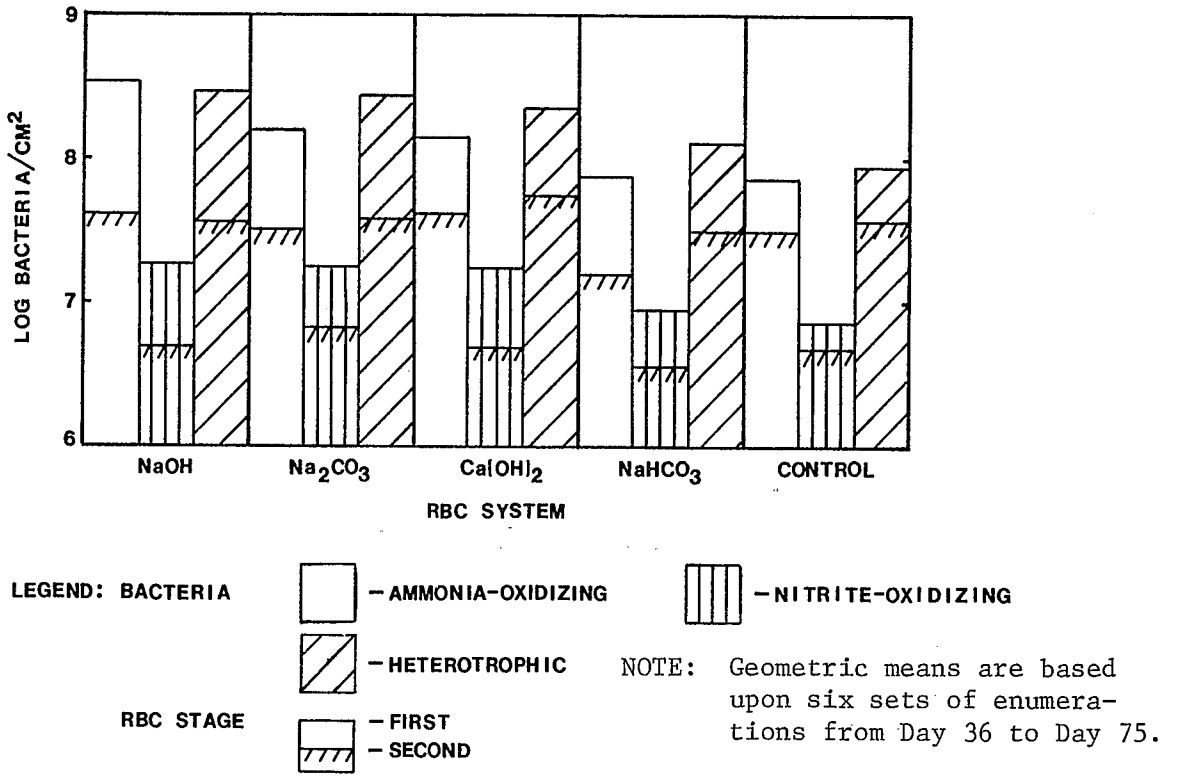


Figure 8. Relative Geometric Mean Populations of Ammonia-Oxidizing, Nitrite-Oxidizing, and Heterotrophic Bacteria on a Unit Disc Area Basis for the Stages of the RBC Systems of the Alkaline Chemical Addition Study

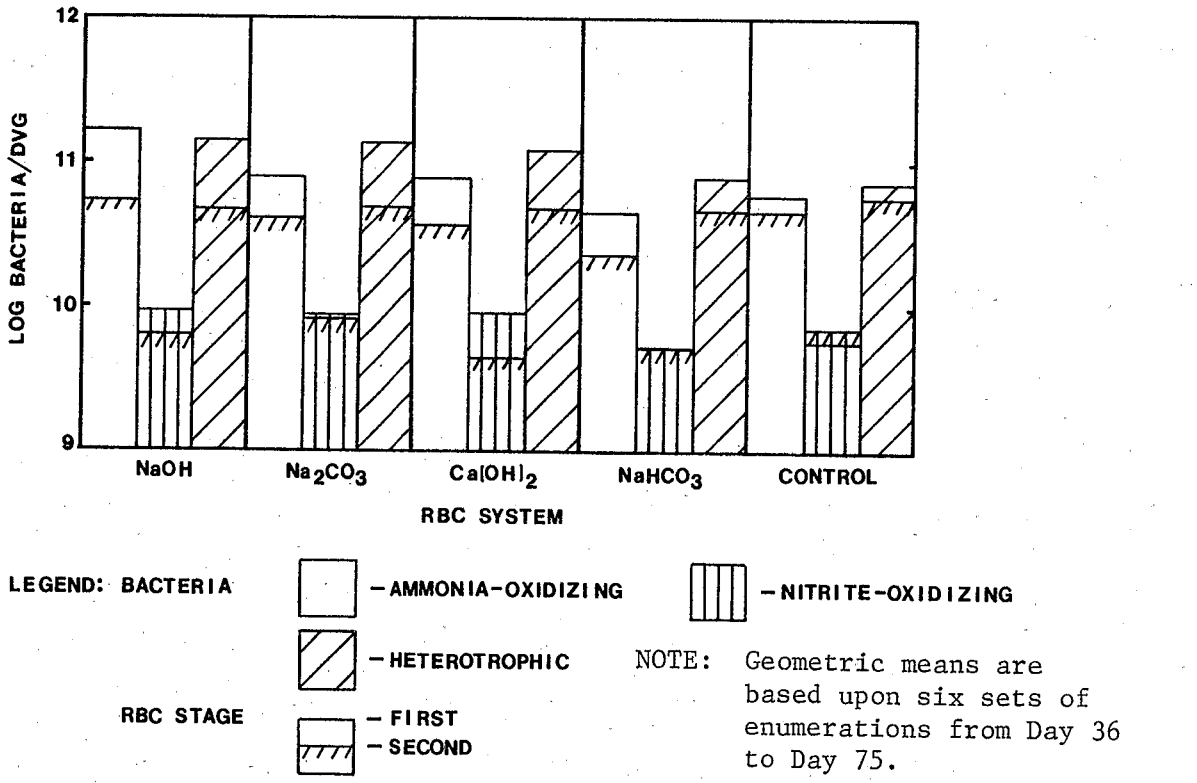


Figure 9. Relative Geometric Mean Populations of Ammonia-Oxidizing, Nitrite-Oxidizing, and Heterotrophic Bacteria on a Unit Weight of Dry Volatile Biofilm Basis for the Stages of the RBC Systems of the Alkaline Chemical Addition Study

Table 12. Ratio of Heterotrophic:Ammonia-Oxidizing: Nitrite-Oxidizing Bacteria for the RBC Stages of the Alkaline Chemical Addition Study[a]

RBC	Heterotrophs:Ammonia-Oxidizers:Nitrite-Oxidizers Stage 1	Stage 2
NaOH	14 : 16 : 1	7.1 : 8.0 : 1
Na_2CO_3	14 : 8.6 : 1	5.5 : 4.7 : 1
$Ca(OH)_2$	12 : 7.8 : 1	11. : 7.9 : 1
$NaHCO_3$	13 : 7.5 : 1	8.1 : 4.0 : 1
Control	11 : 9.4 : 1	7.3 : 6.1 : 1

[a]Ratios are based upon the geometric mean of 6 sets of samples taken at weekly intervals from Day 36 to Day 74.

A summary of the alkalinity destruction rates based upon data obtained from both continuous and batch operations is presented in Table 13. Except for the rate observed in the first stage of the calcium hydroxide RBC, all the alkalinity destruction rates were in the range of commonly accepted values. The unusually low alkalinity destruction rate observed in the bulk solution of the first stage of the calcium hydroxide RBC is attributed to the buildup of calcium carbonate within the biofilm which effectively neutralized some of the acid generated by the nitrifying bacteria within the biofilm. The net result was to reduce the overall amount of alkalinity destroyed in the bulk solution during the nitrification process.

SUMMARY AND CONCLUSIONS

This research examined the short and long-term effect of pH upon the nitrification of wastewater within RBC fixed film systems. In the long-term, the rate of nitrification

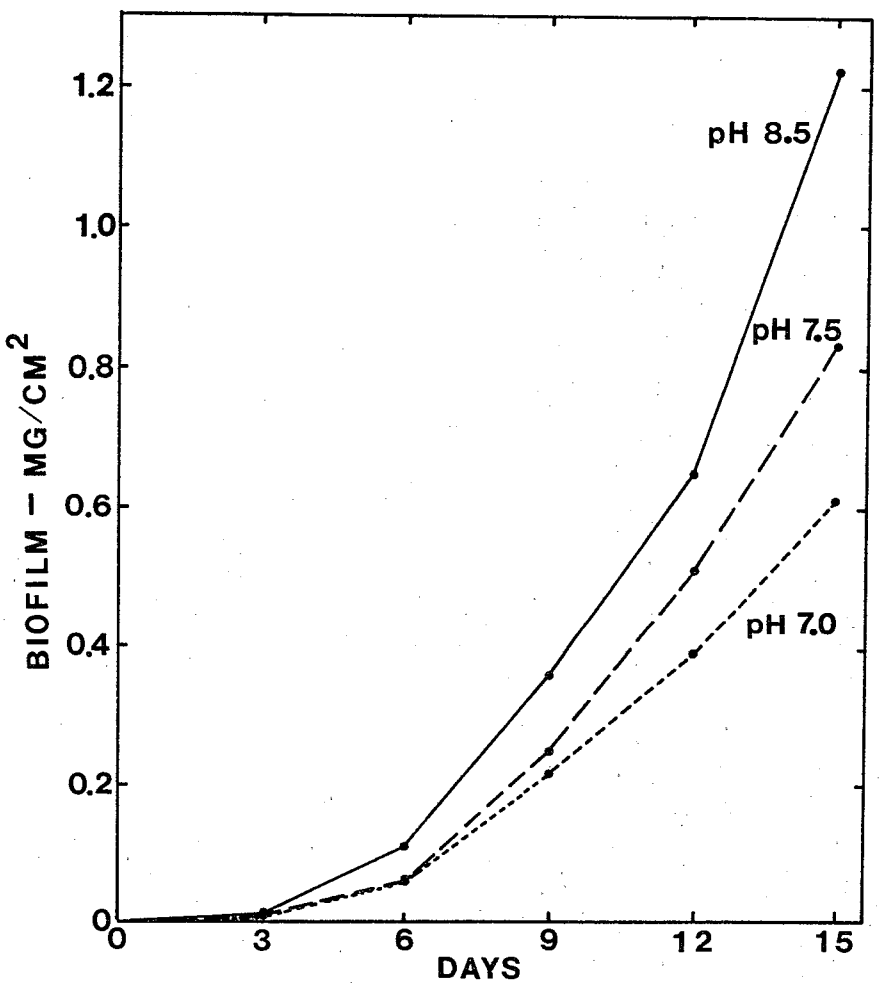


Figure 10. Relative RBC Biofilm Development under pH Conditions from pH 7.0 to pH 8.5

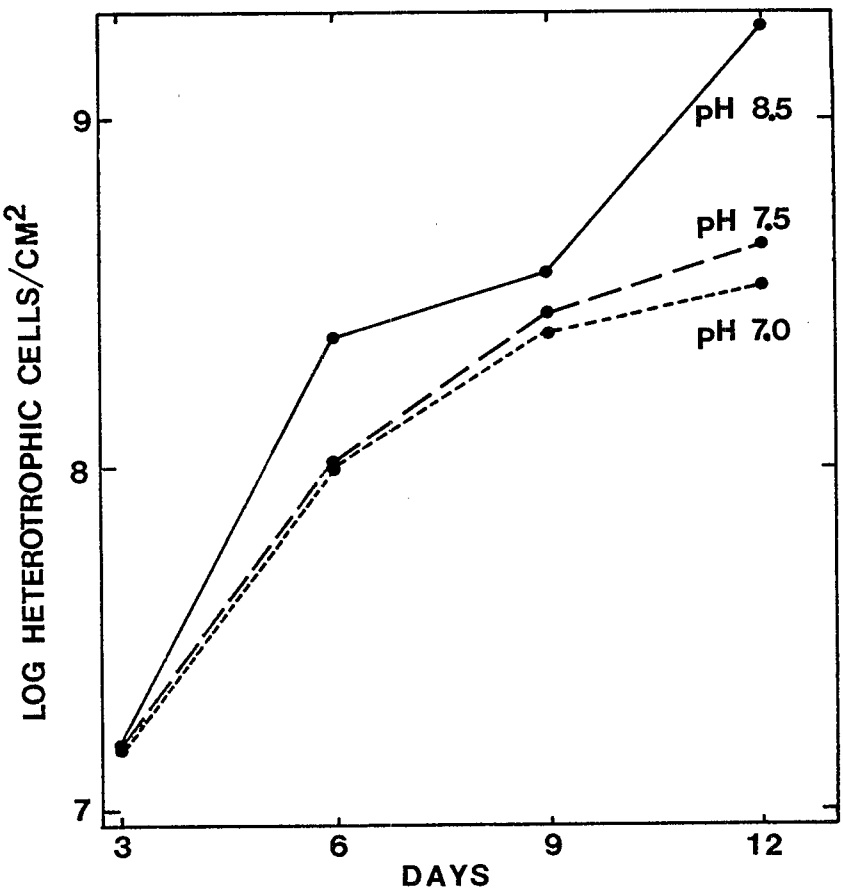


Figure 11. Relative RBC Heterotrophic Bacteria Growth Under pH Conditions from pH 7.0 to pH 8.5

Table 13. RBC Nitrification and Alkalinity Destruction During the Alkaline Chemical Addition Study

RBC - Stage	Alkaline Chemical	Alkalinity Destruction (mg $CaCO_3$/mg NH_3-N) Continuous Operation[a]	Batch Operation
1-1	NaOH	--	6.2
2-1	$NaHCO_3$	--	6.8
3-1	Na_2CO_3	--	7.4
4-1	$Ca(OH)_2$	--	3.8
5-1	Control	7.4	6.6
1-2	--	6.1	--
2-2	--	7.7	--
3-2	--	7.9	--
4-2	--	7.0	--
5-2	--	7.2	--

[a]Based upon data for the continuous operation from Day 38 to Day 75.

within an RBC fixed film system was dependent upon pH. The rate of nitrification increased with increasing pH up to a maximum at pH 8.5. Approximately five weeks of operation were required to clearly observe these differences. The response of a nitrifying RBC system to short-term changes in pH was relatively constant from pH 7.0 to pH 8.5. Below pH 7.0, the adverse effect of pH becomes more pronounced.

However, the absolute level of nitrification was dictated by the biofilm's previous history of nitrification performance. RBC systems continued to nitrify at a relatively high rate after the pH had been reduced suddenly. The nitrogen balances for the various research phases revealed that the relative amount of nitrogen recovered for each RBC system generally was slightly less for the higher pH systems. This result was attributed to low level ammonia stripping as well as the loss of nitrate due to denitrification within the biofilm.

There was no significant difference in the performance of the 2-stage nitrifying RBC systems which received calcium hydroxide, sodium carbonate and sodium hydroxide. The performance levels of the sodium bicarbonate and the control RBC systems were 6 and 16 percent less, respectively, than those of the other three systems. The use of alkaline chemicals to maintain approximately pH 8.5 in the first stage of a 2-stage nitrifying RBC resulted in the removal of approximately 19 percent more ammonia than in the control RBC system. Except for the first stage of the RBC receiving calcium hydroxide for pH adjustment, the range of alkalinity destruction for all RBC systems in the alkaline chemical addition study was from 6.2 to 7.9 mg $CaCO_3$/mg NH_3-N. This result was attributed to a neutralization capacity which developed within the RBC biofilm due to the entrainment of $CaCO_3$. The production of significant amounts of inert material and suspended solids when calcium hydroxide is used, favors the use of sodium carbonate and sodium hydroxide when the nitrification is not followed by secondary clarification.

Higher levels of nitrification for the RBC systems were associated with greater disc biofilm uniformity. In all cases, except for the pH 8.8 RBC of the high pH study, the higher pH RBC systems maintained greater concentrations of volatile biofilm per unit of RBC disc area. The loss of biofilm from the RBC disc surface did not follow the traditionally accepted sloughing pattern. Biofilm did not slough from the disc surface outward. The dominant pattern of biofilm loss was from the biofilm surface inward. This loss was due to hydraulic shear at the biofilm surface. The RBC disc biofilm characteristics changed with time. The initial biofilm was uniform in texture and tan to bronze in color. The biofilm went through an aging process wherein the biofilm became darker and the texture became less uniform; the lower the pH, the less uniform the biofilm. The disc biofilm was affected greatly by low level changes

in CBOD. The maximum rates of nitrification for individual RBC stages were not associated with the maximum biofilm concentrations on the discs. Disc biofilm continued to develop after the individual RBC stages achieved their maximum rate of nitrification. The elevated pH RBC biofilms, which had enhanced nitrification capacities, had higher nitrifying bacterial populations than the lower pH RBC biofilms. The ammonia-oxidizing bacteria generally were favored over the nitrite-oxidizing bacteria with respect to increasing pH. Greater heterotrophic growth and more rapid biofilm development was observed to occur at elevated pH levels.

ACKNOWLEDGEMENT

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SIMULTANEOUS NITRIFICATION AND DENITRIFICATION IN A ROTATING BIOLOGICAL CONTACTOR

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INTRODUCTION

In a Rotating Biological Contactor (RBC) process using a fixed biological film, oxygen gas seldom penetrates into the deepest part of the biofilm. Therefore, denitrifying bacteria usually exist in the anaerobic and inner portions within the biofilm of the RBC process. Denitrifying bacteria existing in the inner anaerobic portion within the biofilm utilize organic matter in the waste water as a source of organic carbon. Nitrite or nitrate nitrogen produced within the aerobic biofilm is partially converted to gaseous nitrogen (N_2 or N_2O) by the denitrifying bacteria.

This phenomenon of simultaneous nitrification and denitrification (referred to as SND) sometimes undoubtedly occurs in the RBC nitrification process. The authors have already observed and reported this phenomenon in RBC pilot plants as well as in a proto-type RBC system(1,2,3). However, it has not yet been verified whether or not gaseous nitrogen is produced in the biofilm. Therefore, the authors have carried out a series of experiments using a completely closed RBC unit to investigate SND. The experimental variables were ammonia loading, organic loading, ammonia concentration and mean resi-

dence time. In this paper, the experimental results concerning SND in an RBC are presented and discussed.

MATERIALS AND METHODS

The experimental apparatus consisted of closed-type reactor and disks made of waterproof veneer boards. Fig.1 shows the single-stage RBC unit used in this experiment. The residence time distribution of water in the reactor without a biofilm perfectly coincided with that of a single completely mixed-flow reactor. In order to develope nitrifying bacteria on the disk surface, artificial waste water (Table.1) containing ammonia and inorganic carbon was fed into the RBC unit. Water temperature and pH were fixed at 30°C and 8.0, respectively. When the nitrifying biofilm developed, the artificial waste water containing methanol as a carbon source for the denitrifying bacteria was added. After a week, the outer layer of the biofilm consisted of heterotrophic bacteria, and a biofilm consisting of the heterotrophic bacteria layer, a nitrifying bacteria layer, and a denitrifying bacteria layer was formed.

At this point, an experiment was started to measure the concentration of inorganic nitrogen (NO_3-N, NO_2-N and NH_3-N) in the effluent and the composition of the gas in the air phase. After one Run was completed, the heterotrophic bacteria layer was washed out by water jet. Then the same procedure was repeated by adding the artificial waste water, depending on the experimental conditions shown in Table.2. The nitrogen removal rate due to simaltaneous nitrification and denitrification includes nitrogen utilized for the cell synthesis of all bacteria concerned in the reaction.

RESULTS AND DISCUSSION

Conposition of the Gas in the Air Phase

Fig. 2(a), (b), (c) show the relationship between the amount of gas in the air phase and the elapsed time after the vent holes were closed. Fig. 2(a) shows the relationship between the reduction rate of oxygen gas in the air phase and the concentration of ammonia in the bulk water. The reduction rate of oxygen gas was influenced by organic loading.

In Run 3-1, when the amount of oxygen in the air phase decreased to 500cc, ammonia appeared in the effulent water. In this case, it is believed that the oxygen gas was not sufficiently supplied to the nitrifying bacterial film, because

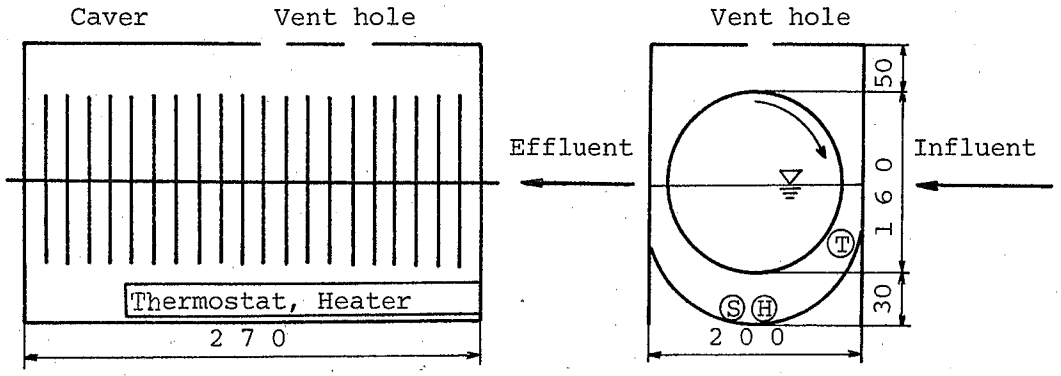
Table. 1 composition of
artificial substrates

Conposition	Conc. (mg/l)
NH_3Cl	382
$NaHCO_3$	1200
NaCl	146
$MgSO_4$ $7H_2O$	123
KH_2PO_4	68

(When ammonia Conc. is 100 mg/l)

Table. 2 Experimental conditions

Run No	Flow rate (cc/min)	NH_3-N loading (g/m^2d)	Organic loading (g/m^2d)	NH_3-N Conc. (mg/l)	M R T (hr)
1 - 1	5	1	3.5	100	13.0
1 - 2	5	1	7.0	100	13.0
1 - 3	5	1	9.0	100	13.0
2 - 1	12	1	4.9	50	5.5
2 - 2	12	1	7.3	50	5.5
2 - 3	12	1	14.7	50	5.5
3 - 1	24	1	6.0	25	2.8
3 - 2	24	1	8.0	25	2.8
3 - 3	24	1	14.0	25	2.8
4 - 1	48	1	5.0	12.5	1.4
4 - 2	48	1	8.3	12.5	1.4
4 - 3	48	1	11.3	12.5	1.4
5 - 1	10	2	8.3	100	6.7
5 - 2	10	2	14.0	100	6.7
6 - 1	20	4	14.6	100	3.3
6 - 2	20	4	26.0	100	3.3



Disk	number	: 20	Air section	: 7.5 l
Disk	diameter	: 16 cm	Liquid volume	: 4.0 l
Disk	thickness	: 0.5 cm	Thermometer	: T
Disk	space	: 1.0 cm	Heater	: H
Disk	area	: 0.82 m^2	Thermostat	: S

Fig. 1 Schematic representation of experimental unit

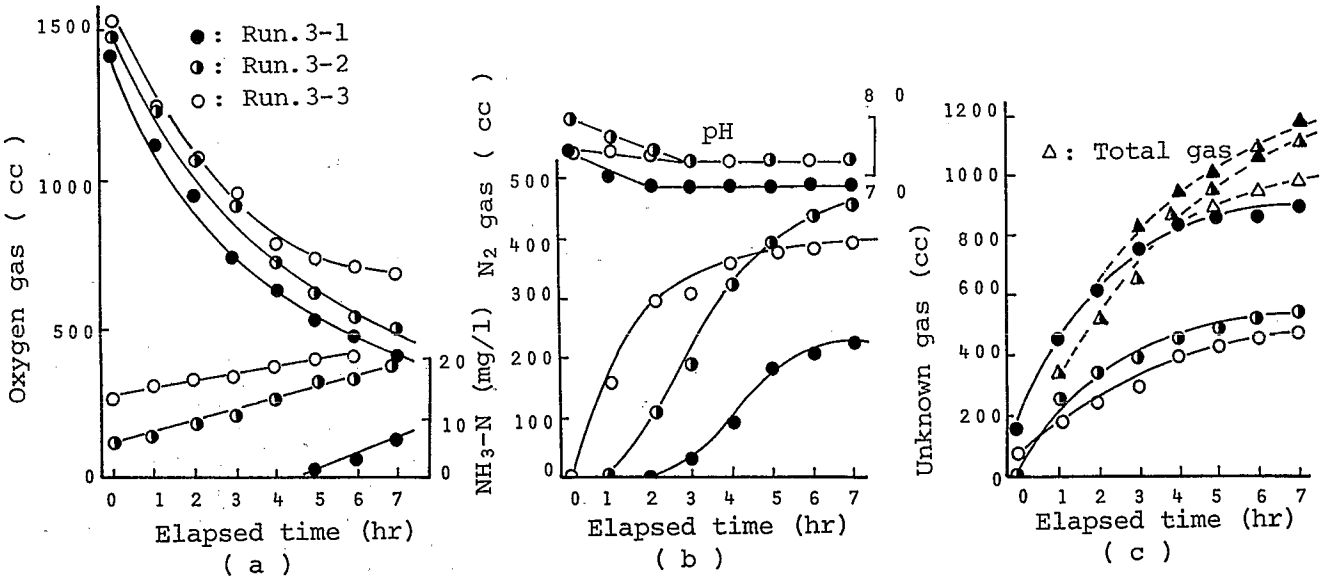


Fig. 2 The relationship between the gas or reduction rate and the elapsed time after the vent holes were clsed.

of the reduction in the partial pressure of oxygen. As the organic loading increased in Runs 3-2 and 3-3, the oxygen fed to nitrification was not considered to be enough even when the vent holes were opened. Fig. 2(b) shows the relationship between the cumulative amount of nitrogen gas and elapsed time.

Fig. 2(c) shows the relationship between elapsed time and the cumulative amount of unknown gas, which could be nitrous oxide (N_2O). The cumulative amounts of the total gas (sum of nitrogen and unknown gas) were almost equal, but the composition of the gas was different in each Run. R.N.Dawson and K.L.Marphy (4) argued that elemental nitrogen was the end product of denitrification above a pH of 7.3, while below a pH of 7.3 nitrous oxide production increases. As pH was less than 7.0 in Run 3-1, it seems that the predominant gas was nitrous oxide.

Changes in the Water Quality

Fig. 3(a), (b), (c) show the relationship between inorganic nitrogens and elapsed time. Fig 3(a) shows the changes in the water quality at an ammonia loadings of 1 g/m^2d and an organic loading of 3.5 g/m^2d. The concentration of nitrate decreased with the elapsed time. On the other hand, the rate of simultaneous nitrification and denitrification increased with the elapsed time. It seems oxygen fed into the biofilm for nitrification became insufficient to accomplish complete nitrification. The relationship between the partial pressure of oxygen in the air phase and the nitrogen removal rate due to SND under the same conditions is shown in Fig. 4(a).

Fig. 3(b) shows the relationship at an ammonia loading of 2 g/m^2d and an organic loading of 8.3 g/m^2d. The concentration of ammonia in the bulk water increased linearly with the elapsed time. Before the experiment began (i.e., during the time the vent holes were open), The partial pressure of oxygen in the gas phase was 21%. Enough oxygen gas was supplied to the biofilm to accomplish complete nitrification. As time elapsed, the nitrogen removal rate decreased, because there was a shortage of oxygen in the biofilm for nitrification.

Fig. 3(c) shows the same relationship at an ammonia loading of 4.0 g/m^2d and an organic loading of 14.6 g/m^2d. The bulk ammonia concentration increased with the elapsed time, because the biofilm for organic oxidation became thicker compered with that shown in Fig. 3(b). When the MRT was long enough (i.e. 13 hr) and the concentrations of ammonia and organic matter were low, the nitrogen removal due to SND in-

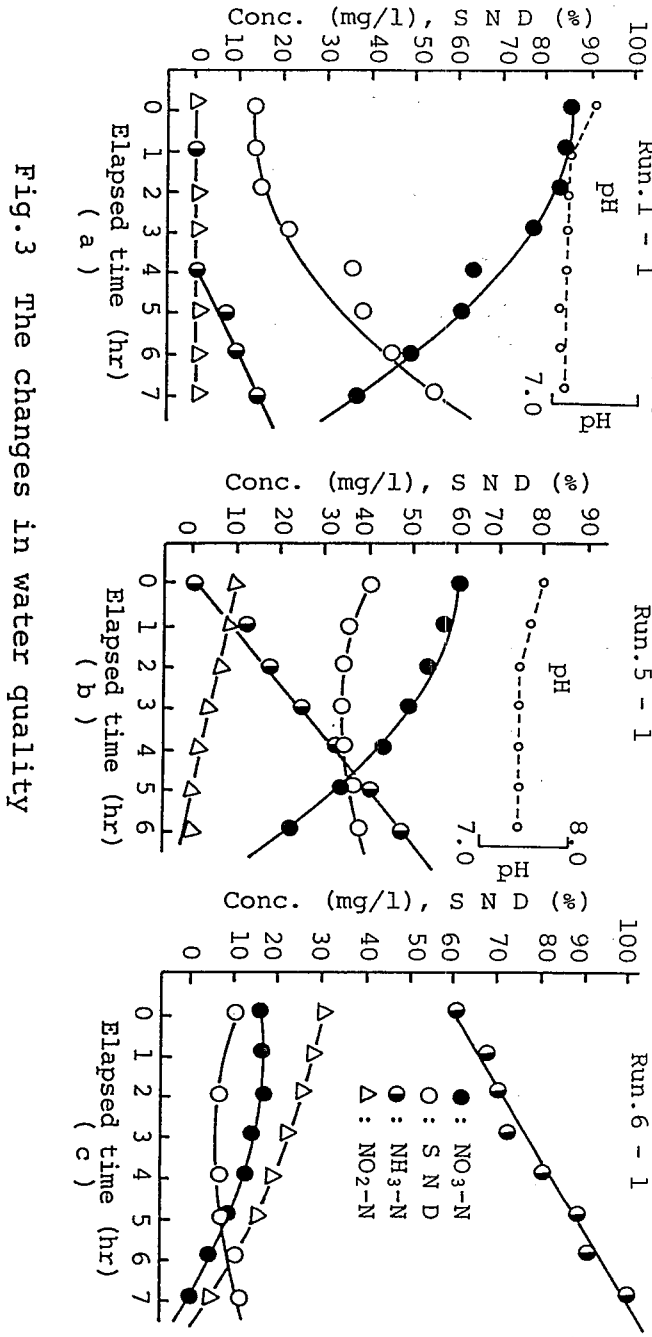


Fig.3 The changes in water quality

creased with the elapsed time. On the other hand, when the concentrations of ammonia and organic matter were high and the MRT was rather short, no change could be seen in the nitrogen removal rate due to SND. The removal rate of nitrogen due to SND depended upon ammonia loading, organic loading, MRT, and the pressure of oxygen in the gas phase.

The Relationship Between a Partial Pressure of Oxygen and SND

Fig. 4 shows the relationship between a partial pressure of oxygen in the gas phase and the nitrgen removal rate due to SND. The experimental results shown in Fig. 4 can be qualitatively explained by the biofilm models shown in Fig. 5. The biofilm consists of a heterotrophic bacteria film for organic oxidation, an autotrophic bacteria film for nitrification, and an anaerobic bacteria film for denitrification. An aerobic biofilm can be considered to be much thicker than the biofilm for nitrification or organic oxidation.

In the case of the attached biofilm shown in Fig. 5(b), the biofilm for organic oxidation was so thick that both the organic matter and DO were mostly consumed within it. Then the biofilm dominant for nitrification became very thin. On the other hand, in Fig. 5(c), the biofilm for organic oxidation was not so thick, therfore, DO penetrated more deeply into the biofilm for nitrification. Therefore, the biofilm for nitrification became thicker than that in 5(b).

Fig. 5(c) shows the case in which the biofilm for organic oxidation slightly covered the biofilm for nitrification. In this case, DO completely penetrated into the biofilm for nitrification. Fig. 4(a) shows the case in which bacteria for organic oxidation slightly covered the biofilm for nitrification at an ammonia loading of 1 g/m^2d and an organic loading of 3.5 g/m^2d. The nitrification rate was sharply reduced at these loadings of ammonia and organics, when the partial pressure of oxygen in the air phase reached less than 10%. On the other hand, when the partial pressure of oxygen increased to more than 10%, the nitrification rate became independent of the partial pressure. However, the nitrogen removal rate due to SND increased with the decrease in the partial pressure of oxygen in the air phase. This can be qualitaitively explained by the biofilm model shown in Fig. 5(c). As the partial pressure of oxygen in the air phase decreased, a part of nitrification biofilm became anaerobic.

In the case of Fig. 4(b), at an ammonia loading of 2 g/m^2 d and an organic loading of 8.3 g/m^2d, the biofilm for nitrification became thicker than that in 4(a). At these loadings

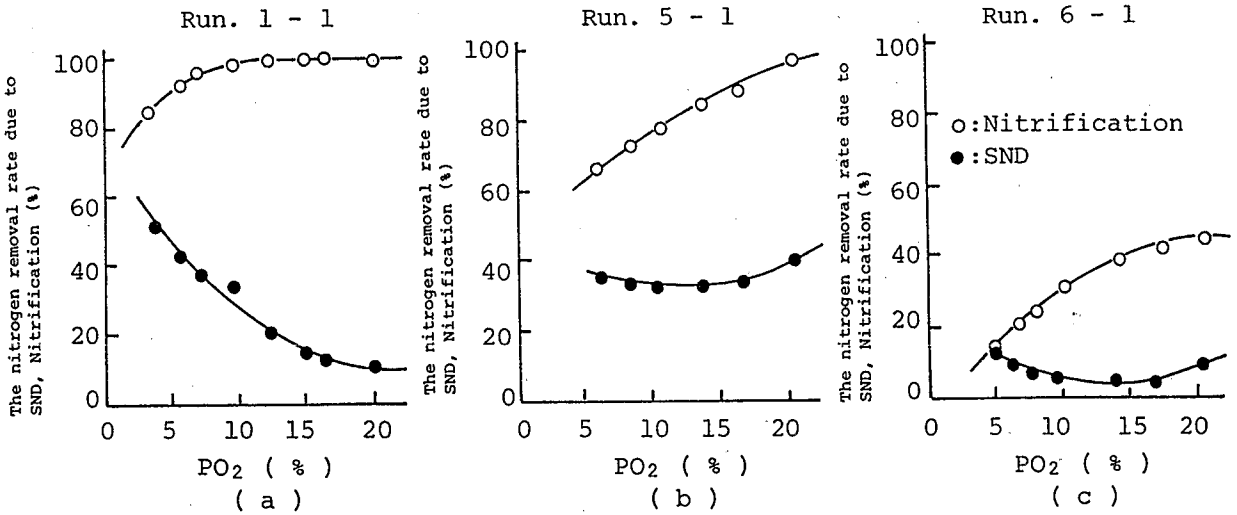


Fig. 4 Relationship between a partial pressure of oxygen in the gas phase and the nitrogen removal rate due to SND

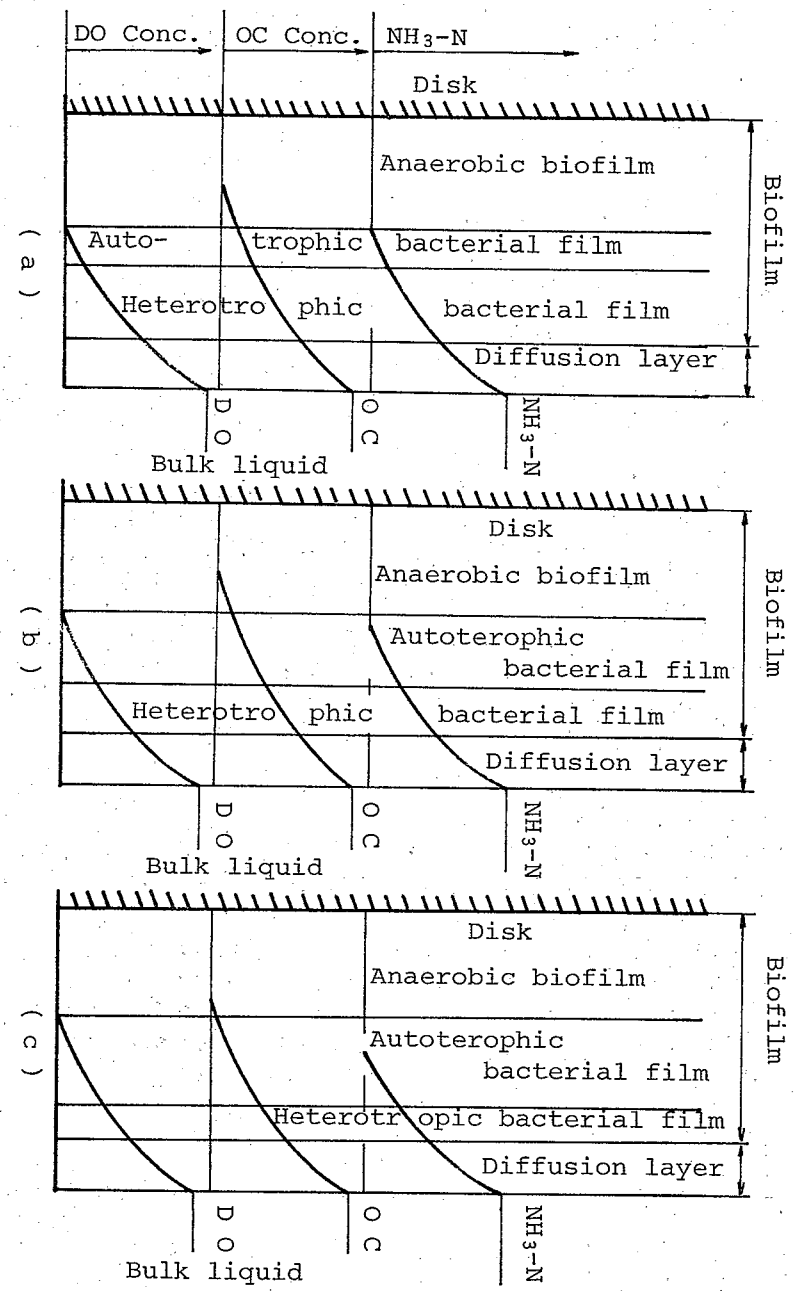


Fig. 5 Biofilm models

of ammonia and organic, the nitrification rate became 100%, when the pressure of oxygen in the air phase equaled 21%. The nitrification rate decreased sharply because the penetration depth of oxygen became shallower as the partial pressure of oxygen in the air phased decreased. This is explained by the biofilm model shown in 5(b). The biofilm for organic oxidation became thicker and the biofilm for nitrification became thinner compared with the biofilm shown in Fig. 5(a), so that the penetration depth of oxygen became shallower than that in 5(a). The partial pressure of oxygen in the air phase decreased, DO did not penetrate deeply into the biofilm for nitrification; and, therefore sufficient nitrification did not occur.

Fig. 4(c) is the case at an ammonia loading of 4.3 g/m^2d and an organic loading of 14.6 g/m^2d. The biofilm attached to the disks for organic oxidation became considerably thicker in comparison with 5(b) or (c), while the biofilm for nitrification became very thin. As a result, the nitrification rate equaled 40%, which was an extremly low value. This experimental result can be explained by the biofilm model shown in Fig. 5(a). Both organic carbon and DO were mostly consumed within the biofilm for organic oxidation, because the high concentration of organic carbon the biofilm for organic oxidation became quite thick. Therefore, the nitrogen removal rate due to SND also decreased, because of the amount of nitrate or nitrite diffusing to the anaerobic biofilm decreased.

Fig. 6 shows one of the experimental results obtained in the batch experiment using heterotrophic bacteria scraped from the outer biofilm layer. This shows that the amount of the nitrifying bacteria contained in the scraped biofilm can be neglected,i.e. only organic oxidation will occuer in the outer biofilm.

Fig. 7(Runs 1,2,3) shows the cases in which MRT and organic loading were changed at a fixed ammonia loading of 1 g/m^2d. Fig. 7(Run 1) shows the case in which two organic loadings of 3.5 and 7.0 g/m^2d were used in the experiment with a fixed influent ammonia concentrationof 100 mg/l and MRT of 13 hrs. The nitrogen removal due to SND became 100%, when the vent holes were opened. It decreased with the elapsed time after the vent holes were closed. The nitrogen removal rate due to SND was sharply reduced, when the partial pressure of oxygen in the gas phase decreased to less than 10%. On the other hand, in the condition in which the partial pressure of oxygen increased to more than 10%, the nitrogen removal rate due to SND became independent of the partial pressure of oxygen.

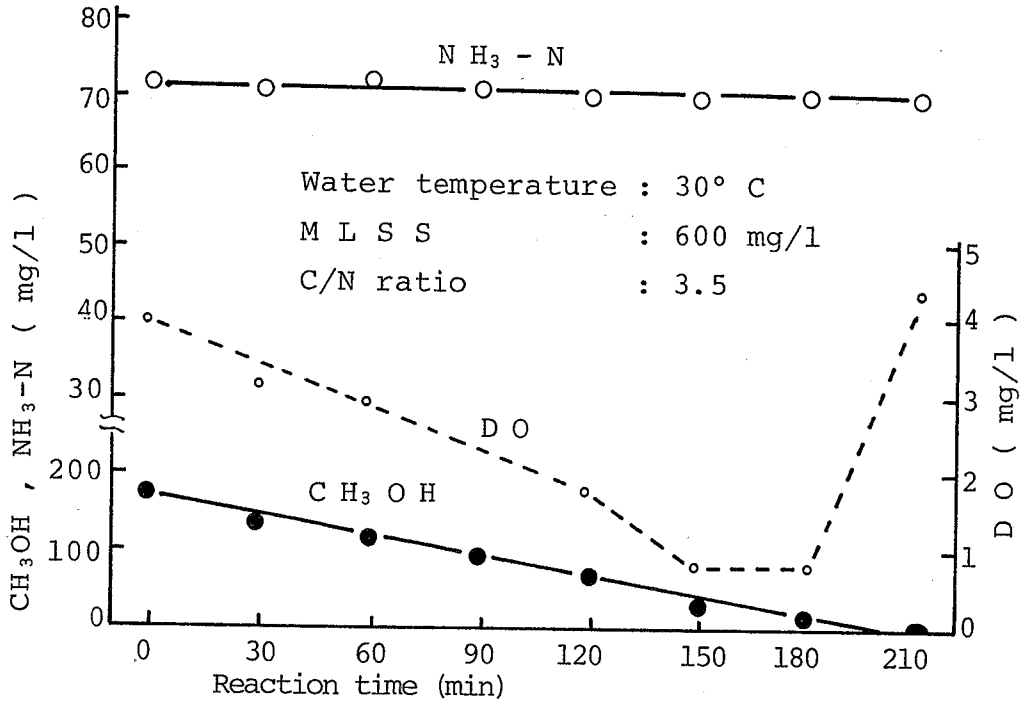


Fig. 6 Ammonia and Methanol Concentrations versus reaction time

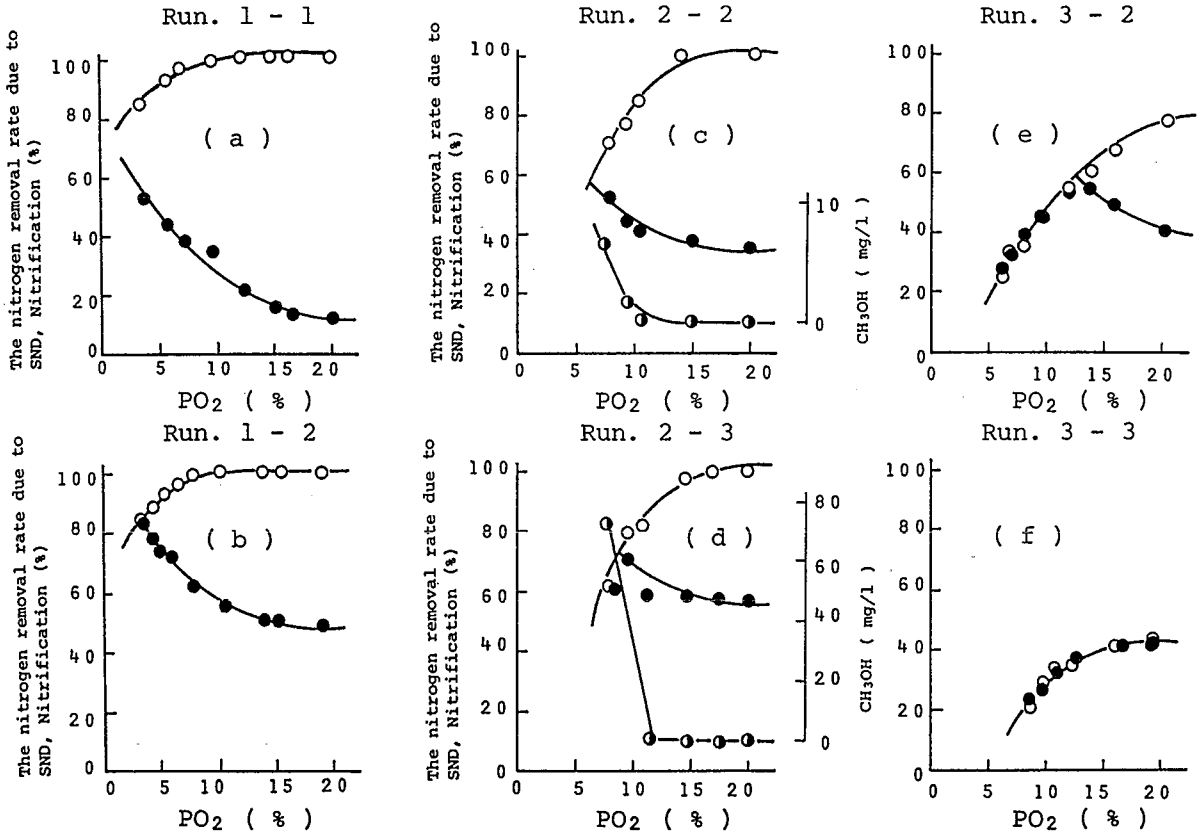


Fig. 7 The pressure of oxygen in the gas phase versus the nitrogen removal rate due to SND, taking MRT and organic loading

In Runs 1-1 and 1-2, it seems that partial oxygen pressure of 10% was enough to acomplish 100% removal. The maximum nitrification rate and SND were obtained, when the partial pressure of oxygen became 3%. Nitrification was rate-limiting when the pressure of oxygen became less than 5%. Where the partial pressure of oxygen increased to more than 10%, the diffusion of organic matter to the anaerobic biofilm was rate-limiting (organic matter limitation). On the other hand, when the pressure of oxygen in the air phase decreased to less than 5%, the diffusion of oxygen to the nitrifying bacterial film was rate-limiting (nitrification limitation). The pattern shown in Fig. 7(a), (b), (c) will be referred to as Pattern A in this paper.

Fig. 7(Run 2) shows the case in which two organic loadings of 7.0 and 10 g/m^2d were used with fixed conditions of an ammonia concentration of 50 mg/l and a MRT of 5.5 hrs. Run 2-1 also adhered to Pattern A. When the pressure of oxygen in the gas phase became 10%, organic matter appeared in the effluent water. In Run 3-3, organic matter in the effluent water did not appear until the pressure of oxygen reached 12%, but the nitrogen removal rate due to SND increased sharply below a pressure 8%. SND changed from organic matter limitation to nitrification limitation at a partial oxygen pressure of 10%. Nitrification and SND showed the same decreasing pattern, when the partial pressure of oxygen in the air phase became less than 10%. The pattern shows in Fig 7. (d), (e) will be referred to as pattern B.

Fig. 7(Run 3) shows the case in which two organic loadings of 8.0 and 10 g/m^2d were used with a fixed condition of an ammonia concentration of 25 mg/l and a MRT of 2.8 hrs. In Run 3-1, on condition that the pressure of oxygen in the air phase reached 20%, SND became organic matter limitation. On the other hand, when the partial pressure of oxygen in the air phase decreased to less than 20%, it became nitrification limitation. In Run 3-2, nitrification and SND show the same decreasing pattern. The pattern shown in Fig. 7(f) will be referred to as Pattern C. As explained above, the patterns of SND can be classified into three types depending on the experimental conditions.

The Relationship between the Concentration Ratio of Methanol to Ammonia and SND

The relationship between the concentration ratio of methanol to ammonia (i.e., the C/N ratio) and the nitrogen removal rate due to SND is shown in Fig 8. In this experiment, orga-

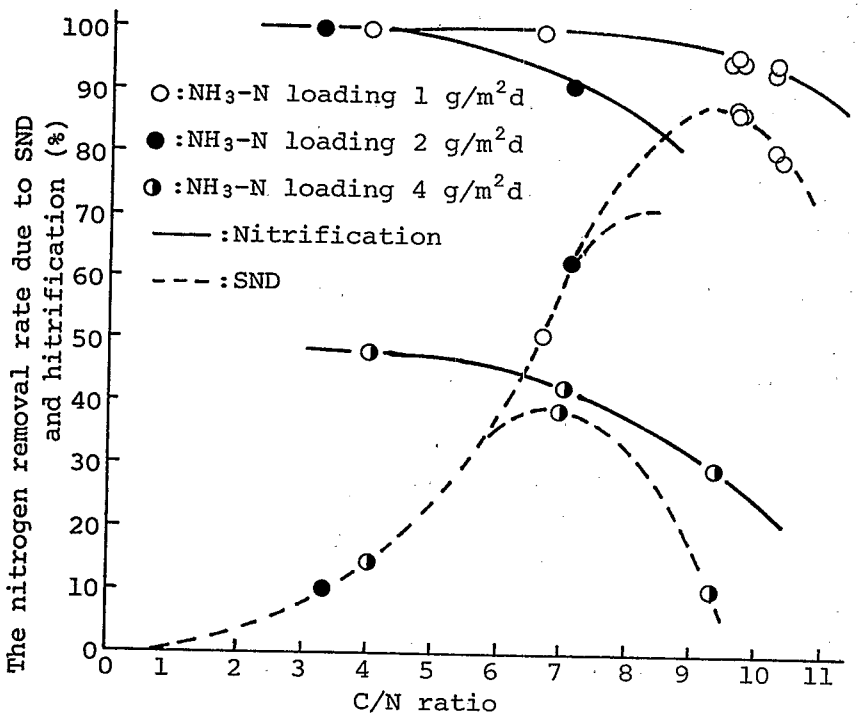


Fig. 8 The nitrogen removal rate due to SND versus C/N ratio

nic loading ranged from 3.5 to 36 g/m^2d. Organic loadings of 3.5, 7, 9, and 10 g/m^2d were used in the experiment with an ammonia loading of 1 g/m^2d. At these loading rates of ammonia and organic matter, a nitrification rate equal to 100% was obtained, even when the C/N ratio was changed from 4 to 7. However, when the C/N ratio became more than 7, the nitrification rate became lower. At the same loading rate, the nitrogen removal rate due to SND increased significantly with an increase in the C/N ratio. The maximum nitrogen removal rate due to SND was ofbtained at a C/N ratio equal to about 9. Organic loadings of 8.3, 14 and 18g/m^2d were used in the experiment at an ammonia loading of 2 g/m^2d. The nitrification rate sharply decreased. The nitrogen removal rate due to SND increased significantly with the increase in the organic loading and the maximum nitrogen removal rate due to SND was obtained at C/N ratio of around 8. Organic loadings of 14.6, 28 and 36 g/m^2d were used in the experiment with a fixed ammonia loading of 4.3 g/m^2d. The nitrification rate decreased with the increase of C/N ratio. The nitrogen removal rate due to SND became very high as the C/N ratio increased and the maximum nitrogen removal rate due to SND was obtained at a C/N ratio of 7.

Fig. 9 and 10 show the relationship between the C/N ratio and SND, when sodium formic acid and ethylene glycol were added as organic carbons. A nirtogen removal rate of 20% due to SND was obtained even at a C/N ratio of zero, because a landfill leachate (Haginodai) was used for the raw waste water. When sodium formic acid was added as an organic carbon, the nitrogen removal rate of 80% due to SND was obtained at a high C/N ratio of 35.

These results demonstrate it was difficult for the heterotrophic and anaerobic bacteria to use the sodium formic acid. On the other hand, in the case where ethylene glycol was added as organic carbon, a nitrogen removal rate of 90% due to SND was obtained at the low C/N ratio of 4. Based on the experimental data, the authors belive the effectiveness of the organic carbon as the carbon source of SND depends on the diffusivity and biodegradibility of the organics which influence the distributions of heterotrophic, nitrifying, and denitrifying bacteria concentrations in the biofilm.

Simultaneous Nitrification and Denitrification in a Proto-type RBC Plant

Since Novenber 1976, a Rotating Biological Contactor has been treating leachate from the Miyazaki Citiy Haginodai Landfill (5). The concentration of total nitrogen in the effluent

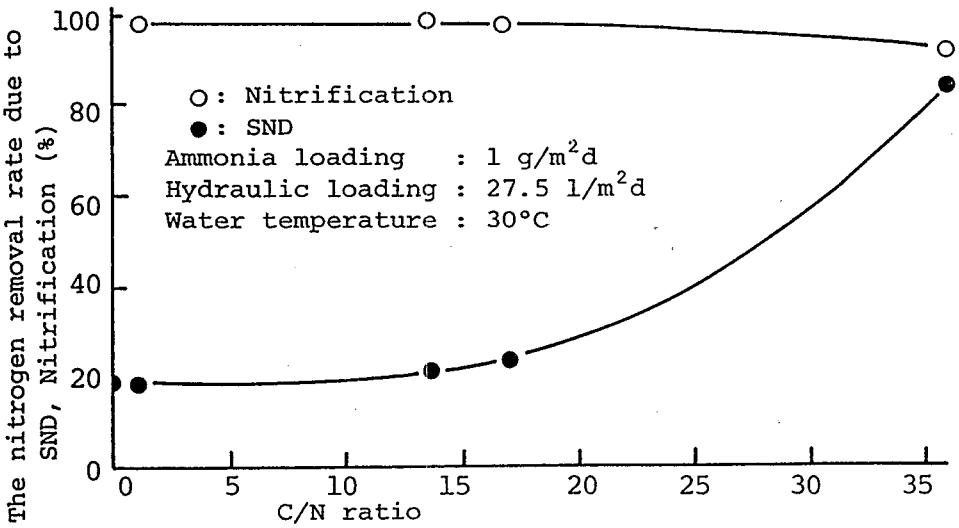


Fig. 9 The nitrogen removal rate due to SND versus the elapsed time (Sodium formic acid as organic carbon)

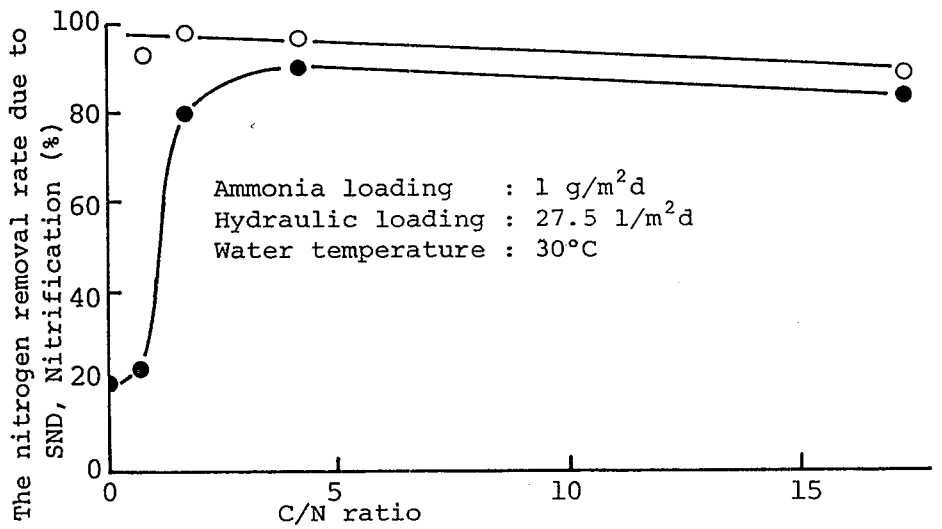


Fig. 10 The nitrogen removal rate due to SND the elapsed time (Ethylene glycol as organic carbon)

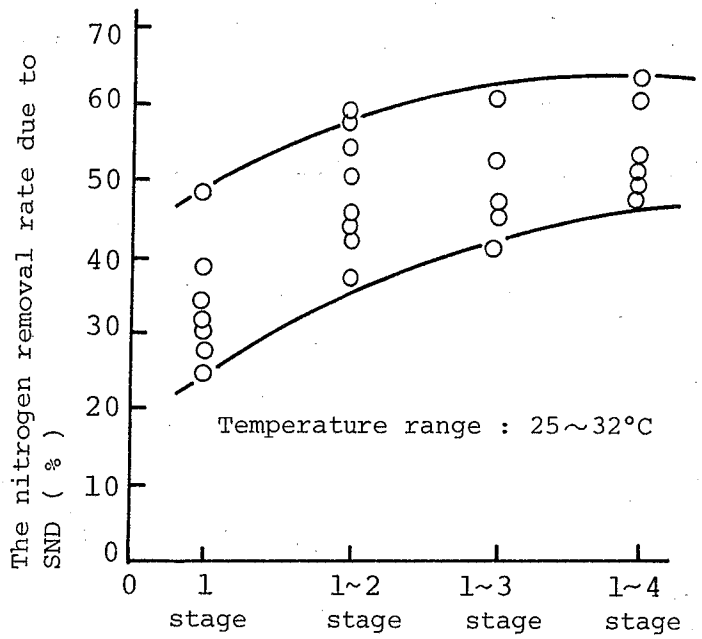


Fig. 11 The removal rates of SND for individual stage in aerobic reactor

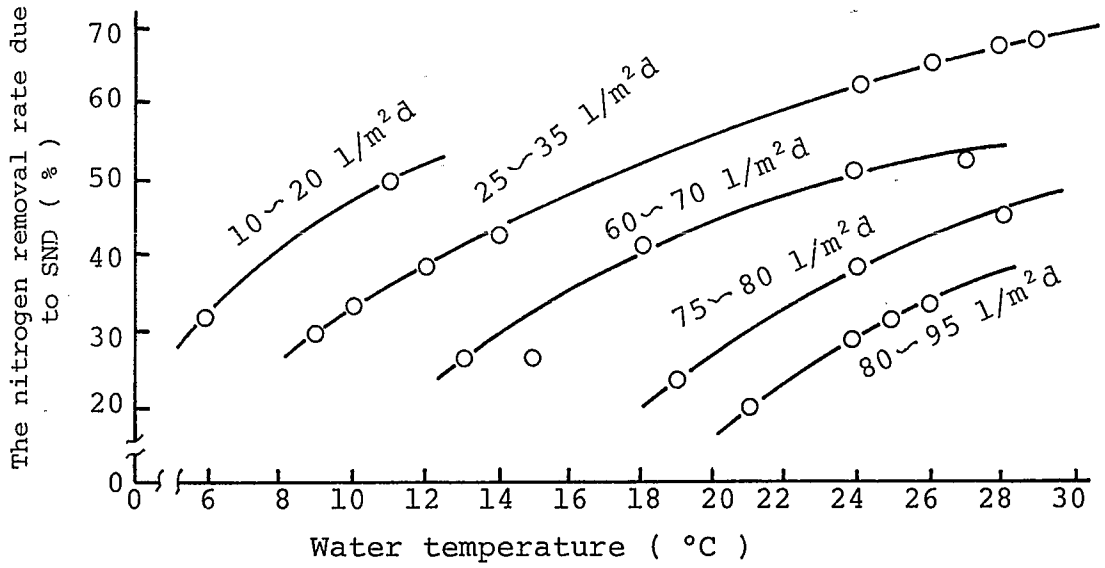


Fig. 12 The water temperature versus SND, taking hydraulic loading as a parameter

water has constantly been less than 10 mg/l. In this experiment, the nitrogen removal rate due to SND was obtained from aerobic RBC.

Fig. 11 shows the cumulative nitrogen removal rate due to SND. The nitrogen removal rate due to SND increased at higher temperatures. SND occurred significantly in the first two stages but hardly occurred at all in the latter stages.

Fig. 12 shows the relationship between SND and water temperature, with hydraulic loading as a parameter. SND became significant at higher temperatures and lower hydraulic loadings. Hydraulic loadings of less than 60 to 70 l/m^2d should be used in this plant to obtain a nitrgen removal rate due to SND greater than 40% at 20 °C.

SUMMARY AND CONCLUSIONS

The phenomenon of SND in a RBC process was confirmed by measuring nitrogen gas production. The SND in a RBC was experimentally studied in terms of mean residence time, organic loading, water temperature, and the partial pressure of oxygen in the air phase. Most of the conventional biological processes for nitrogen removal consist of a series of unit processes which perform organic oxidation, nitrification, and denitrification separately. On the other hand, in the case of nitrogen removal using the RBC nitrification process, all the organic oxidation, nitrification, and denitrification could be accomplished in the same reactor. The obtained results are summarized as follows:

1. In a closed RBC for nitrification, nitrogen gas increased in the air phase, while oxygen gas decreased.
2. The nitrogen removal rate due to SND depended upon the MRT, C/N ratio, water temperature, partial pressure of oxygen, and ammonia concentration. When the other parameters were fixed, the optimum C/N ratio existed for the maximum nitrogen removal rate due to SND.
3. The SND patterns were classified into three types depending on the experimental conditions.

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DENITRIFICATION IN A SUBMERGED BIO-DISC SYSTEM WITH RAW SEWAGE AS CARBON SOURCE

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INTRODUCTION

An investigation of denitrification by biofilms has been performed in two submerged bio-disc units with municipal sewage as carbon source. The project is based upon the process that the authors presented in the Proceedings of The First National Symposium on Rotating Biological Contactor Technology (1). The flow sheet of the process is shown in Figure 1. Nitrate-rich effluent is recycled to the inlet of an anoxic tank where denitrification takes place with raw municipal sewage as carbon source.

In this paper results are presented from a later more thorough study of the denitrification part, in order to establish the basic design criteria for the process. The goal was to find the denitrification rate, - temperature dependency, - pH dependency, - energy consumption and - alkalinity production.

EXPERIMENTAL ARRANGEMENT

The experiments were carried out in two plexi-glass bio-disc units arranged in parallel . The units called RBC A

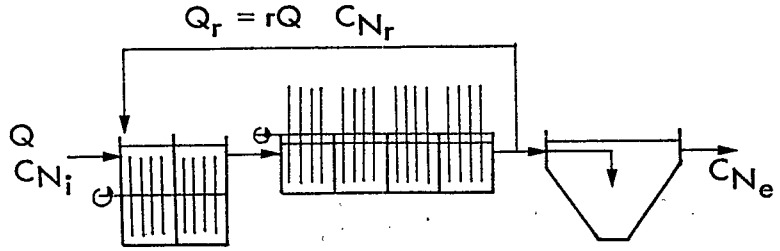

Figure 1. Proposed process for nitrogen removal in biodisc plants.

and RBC B were identical, each containing one 300 mm diameter disc with a total biofilm area of 0.15 m^2. The discs were rotated at 17.5 rev./min., which equals a peripheral velocity of 27.5 cm/sec. The tank volumes were 10.5 litres. The units were equipped with double walls. About 24 hours prior to sampling, the inner walls were removed, scraped, washed and reinstalled. This was done to avoid the wall-growth effects, that otherwise can be significant in pilot-scale plants.

Ahead of each RBC unit a tank for pH and temperature control was installed.

The pH was adjusted by means of an automatic dosing equipment, adding sulfuric acid or sodium hydroxide according to signals given by pH-electrodes. The temperature was adjusted with water from water-baths circulating in copper tubing. In addition the RBC units and the tanks for pH and temperature control were heavily insulated. The flow sheet of the experimental set-up is shown in Figure 2.

The raw sewage was presettled before entering the raw water tanks. Samples were analysed for influent and effluent NO_x -N, NO_2-N, alkalinity and $SBOD_5$. In addition SCOD and DOC were measured during the temperature and pH runs. Flow rates, temperature, pH and dissolved oxygen concentrations were observed for each run. $SBOD_5$ was chosen as the main parameter for measuring the organic content, since it was believed to be the parameter that best describes the organic part available for the denitrifying organisms.

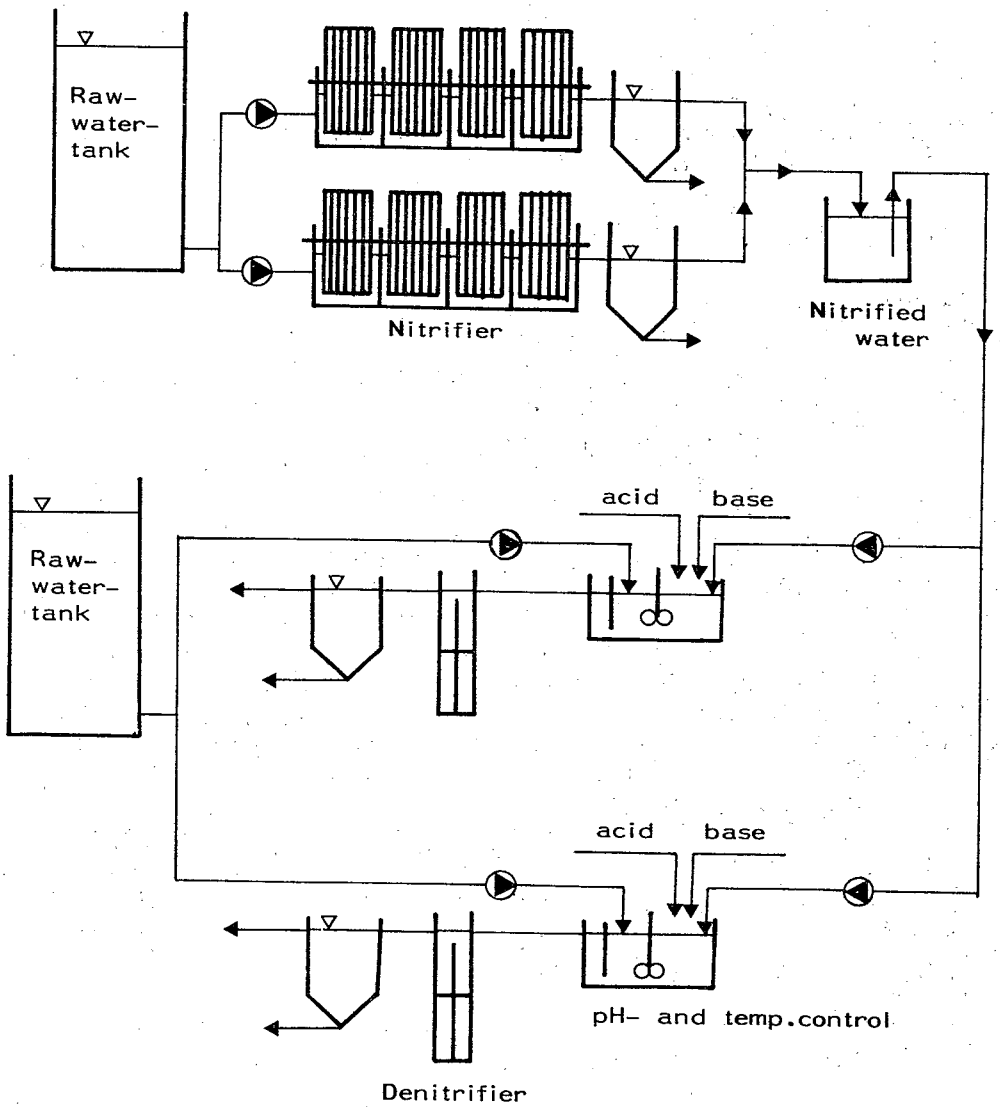

Figure 2. Experimental set-up.

Initially both RBC units were run at 15°C and pH 7. In the temperature- and pH dependency experiments, temperature and pH in RBC A were varied. RBC B acted as a reference unit at constant temperature and pH.

RESULTS AND DISCUSSION

Denitrification rates

Several factors are influencing the denitrification rate, such as oxygen concentration, pH, temperature, carbon source, nitrate loading and organic loading.

Denitrification has to be carried out at anoxic conditions. It seems as if true anaerobic conditions in the liquid is not necessary, and that 1-2 mg O_2/ℓ does not influence denitrification in biofilms (2,3). In this experiment the oxygen concentrations were usually below 0.4 mg O_2/ℓ, and never exceeded 1.0 mg O_2/ℓ . Thus oxygen is not considered to be a limiting factor for the denitrification rates found in this study.

Temperature and pH were held constant. (15°C and pH 7). Various proportions and amounts of raw water and nitrified water were fed to the denitrifiers, in order to cover a broad range of hydraulic, organic and nitrate loadings.

When modelling a denitrification system many researchers use a Monod relation, applying excess methanol and taking nitrate as the limiting substrate. Our experiments showed that the denitrification rates were very dependent upon the $SBOD_5$ concentration. Supposing that the RBC units are complete-mix reactors, the denitrification rates should be a function of effluent NO_x-N and $SBOD_5$ concentrations. Adding a "Monod" term to the steady-state removal equation given by Kornegay and Andrews (4) gives:

$$Q\,(S_{0\ NO_x\text{-}N} - S_{1\ NO_x\text{-}N}) = \frac{\hat{\mu}}{y}(A)(X)(d)\left(\frac{S_{1\ NO_x\text{-}N}}{K_{S\ NO_x\text{-}N} + S_{1\ NO_x\text{-}N}}\right)\left(\frac{S_{1\ SBOD_5}}{K_{S\ SBOD_5} + S_{1\ NO_x\text{-}N}}\right) \quad (1)$$

where Q = hydraulic flow rate (L^3/T)

$S_{0\ NO_x\text{-}N}$, $S_{1\ NO_x\text{-}N}$ = influent and effluent concentration of NO_x-N, respectively (M/L^3)

$S_{1\ SBOD_5}$ = effluent concentration of $SBOD_5$ (M/L^3)

$\hat{\mu}$ = maximum specific growth rate (T^{-1})

y = yield

A = area of biological film (L^2)

X = concentration of organisms in the biological film (M/L^3)

d = thickness of the active biological layer (L)

$K_{S\ NO_x\text{-}N}$, $K_{S\ SBOD_5}$ = saturation coefficients for NO_x-N and $SBOD_5$, respectively (M/L^3)

Assuming that $\hat{\mu}$, Y, X and d are constants, which is only partly true, Equation 1 can be rewritten to give the denitrification rate:

$$R_{NO_x\text{-}N} = C\left(\frac{S_{1\ NO_x\text{-}N}}{K_{S\ NO_x\text{-}N} + S_{1\ NO_x\text{-}N}}\right)\left(\frac{S_{1\ SBOD_5}}{K_{S\ SBOD_5} + S_{1\ SBOD_5}}\right) \quad (2)$$

where $R_{NO_x\text{-}N}$ = denitrification rate (M/L^2T)

C = constant (M/L^2T)

Least square regression was used to find the equation that best described our results (5). A lot of functions, including the one described by Equation 2, were tested. The results for the model with best fit and for the Monod-type model are listed in Table I.

Table I. Denitrification rate at pH 7 and 15°C.

Rate expression , mg/m^2h	Degr. of freedom	Residual mean square	r for R_{NO_x-N} obs. versus R_{NO_x-N} pred.	
$R_{NO_x-N} = 12.72 \left(\frac{L_{NO_x-N}}{335.3 + L_{NO_x-N}}\right) S_{0\ SBOD_5}$ and $R_{NO_x-N} \leq L_{NO_x-N}$	96	4486	0.9166	(3)
$R_{NO_x-N} = 4305\left(\frac{S_{1\,NO_x-N}}{0.035 + S_{1\,NO_x-N}}\right)\left(\frac{S_{1\,SBOD_5}}{279 + S_{1\,SBOD_5}}\right)$	94	13950	0.7216	(2)

where L_{NO_x-N} = NO_x-N load , mg/m^2h

$S_{0\ SBOD_5}$ = influent $SBOD_5$ concentration , mg/l

$S_{1\ SBOD_5}$ = effluent $SBOD_5$ concentration , mg/l

S_{NO_x-N} = effluent NO_x-N concentration , mg/l

Denitrification rates are plotted in Figure 3. Also shown are the rates predicted by Equation 3. Figure 4 and Figure 5 show observed and predicted denitrification rates using "the best fit model" and the Monod-type model, respectively.

Using municipal sewage with variations in both composition and strength, it is not possible to attain steady-state conditions. Influent $SBOD_5$ concentrations covered a range from 9 to 80 mg/ℓ and influent NO_x-N concentrations varied from 5.35 to 20.80 mg/ℓ. In addition we used theoretical hydraulic residence times varying from 0.93 to 3.98 hours, with 1.15 hours as a typical value. As shown in Figure 4, equation 3 describes reasonably well the observed denitrification rates. This means that in our RBC reactors the denitrification was influenced mainly by the influent organic strength ($SBOD_5$) and the nitrate load.

The Monod-type model shows considerably more spread in the denitrification rates (Figure 5). One of the reasons may be that our system was not steady-state. Equation 2 gives us the saturation coefficients, $K_{S\ NO_x\text{-}N} = 0.035$ mg NO_x-N/ℓ and $K_{S\ SBOD_5} = 279$ mg $SBOD_5$/ℓ. For practical purposes this means that denitrification is a 0.order reaction with respect to NO_x-N concentration. Using municipal sewage and the resirculation system proposed in Figure 1, the $SBOD_5$ concentration is unlikely to reach the $K_{S\ SBOD_5}$-value. Denitrification can therefore be considered a 1.order reaction with respect to $SBOD_5$.

Requa and Schroeder (6) found $K_S = 0.06$ mg NO_3-N/ℓ in a biofilm reactor. Murphy et.al (7) found that denitrification in a submerged RBC in the presence of an adequate carbon source was independent of the NO_x-N concentration. This supports our assumption that denitrification is 0.order with respect to NO_x-N concentration.

Temperature dependency

In the temperature and pH dependency experiments, RBC B was used as a reference unit. Variations in the raw water, for instance appearance of toxic material, could otherwise give a wrong picture of the temperature or pH dependency. RBC B was kept constant at 15°C, and the relative denitrification rate at this temperature is put equal to 1.00.

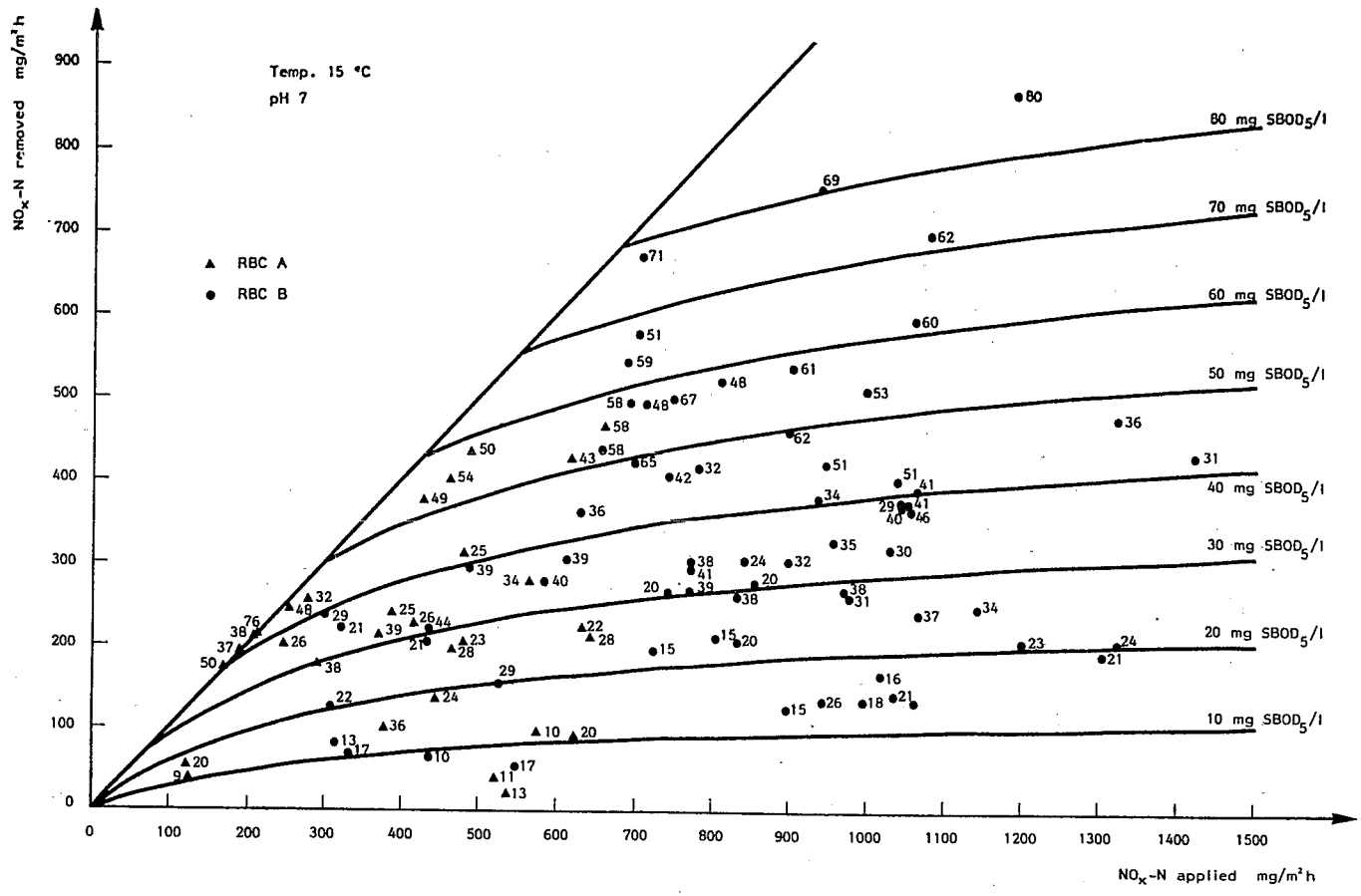

Figure 3. Denitrification rates as a function of applied NO_X-N and influent $SBOD_5$ concentration (indicated by the figures). Solid lines show denitrification rates predicted by Equation 3.

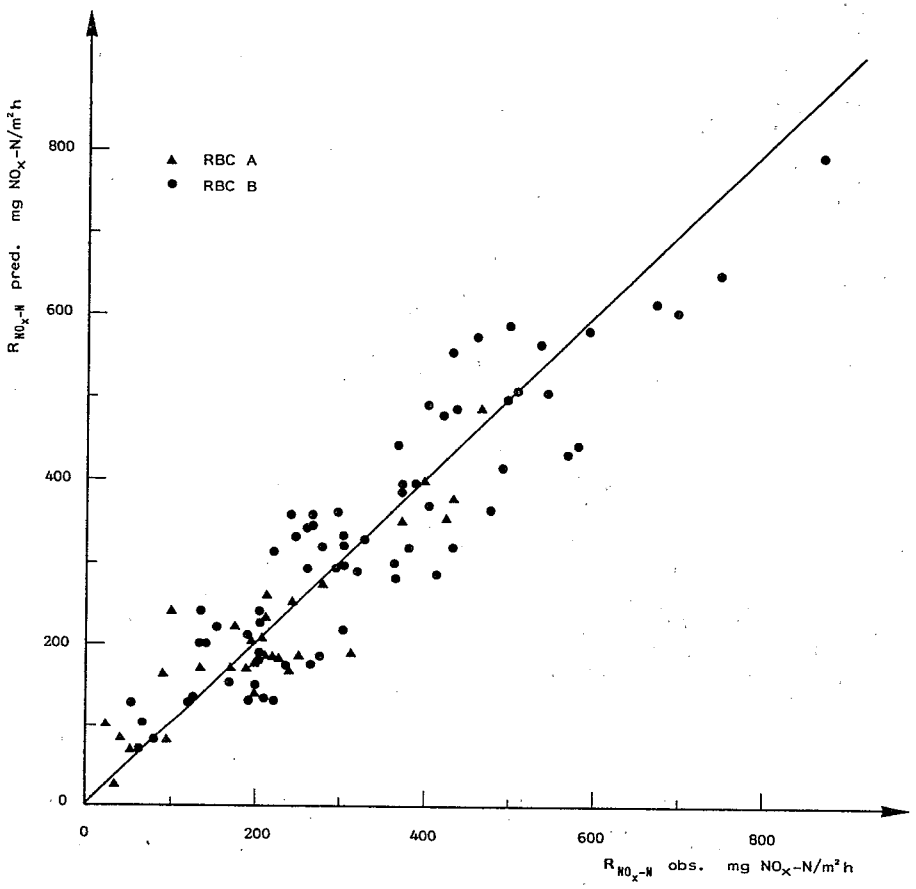

Figure 4. Observed versus predicted denitrification rates for the model with best fit (Equation 3).

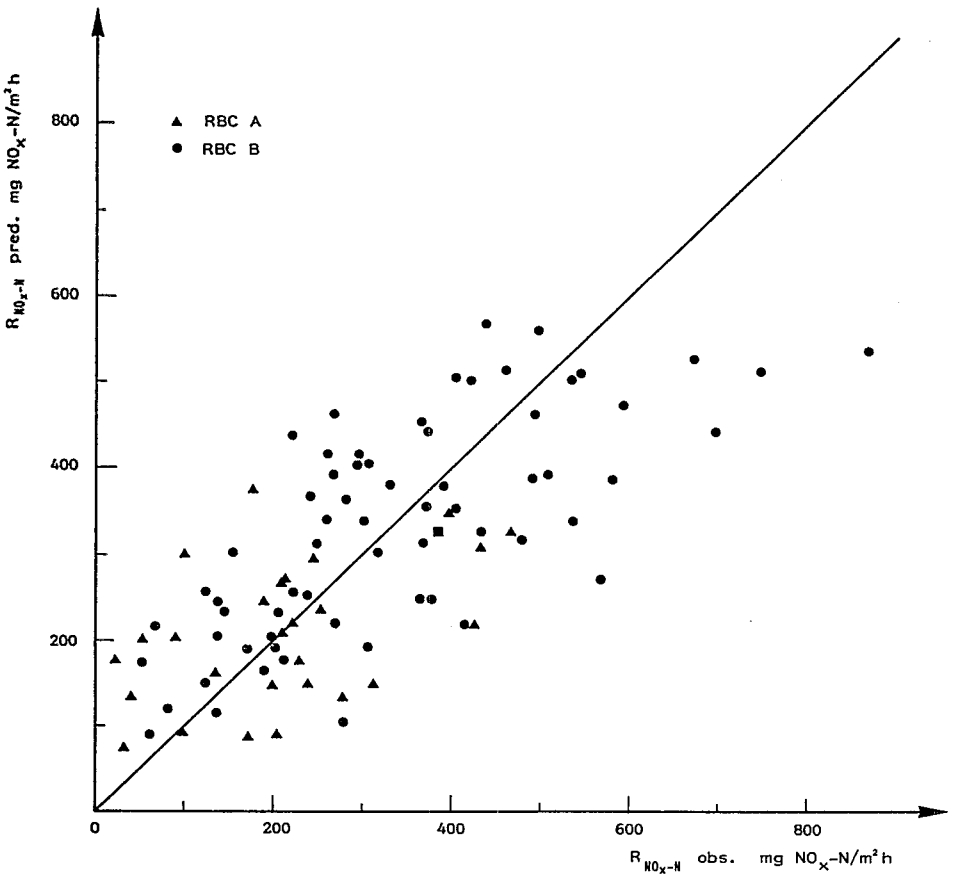

Figure 5. Observed versus predicted denitrification rates for the Monod-type model (Equation 2).

The relative denitrification rates in RBC A were determined according to Equation 4:

$$RD = \frac{\text{"Rate RBC A"}}{\text{"Rate RBC B"}} \qquad (4)$$

The Arrhenius relationship is often used to describe temperature effects:

$$R_T = A \cdot e^{-E/RT} \qquad (5)$$

Equation 5 was rewritten to give an Arrhenius plot:

$$RD = k \cdot e^{-E/RT}$$

$$\ln (RD) = -E \cdot \left(\frac{1}{RT}\right) + \ln k \qquad (6)$$

Using least square regression (Figure 6) we found an energy of activation of 39 670 J/mole. Figure 7 shows the Arrhenius equation and the observed values for the relative denitrification rates.

Temperature dependencies can also be expressed by:

$$Q_{10} = \frac{R_{T+10}}{R_T} \qquad (7)$$

or:

$$R_{T2} = R_{T1} \cdot \Theta^{(T_2 - T_1)} \qquad (8)$$

Results from this study are listed in Table II.

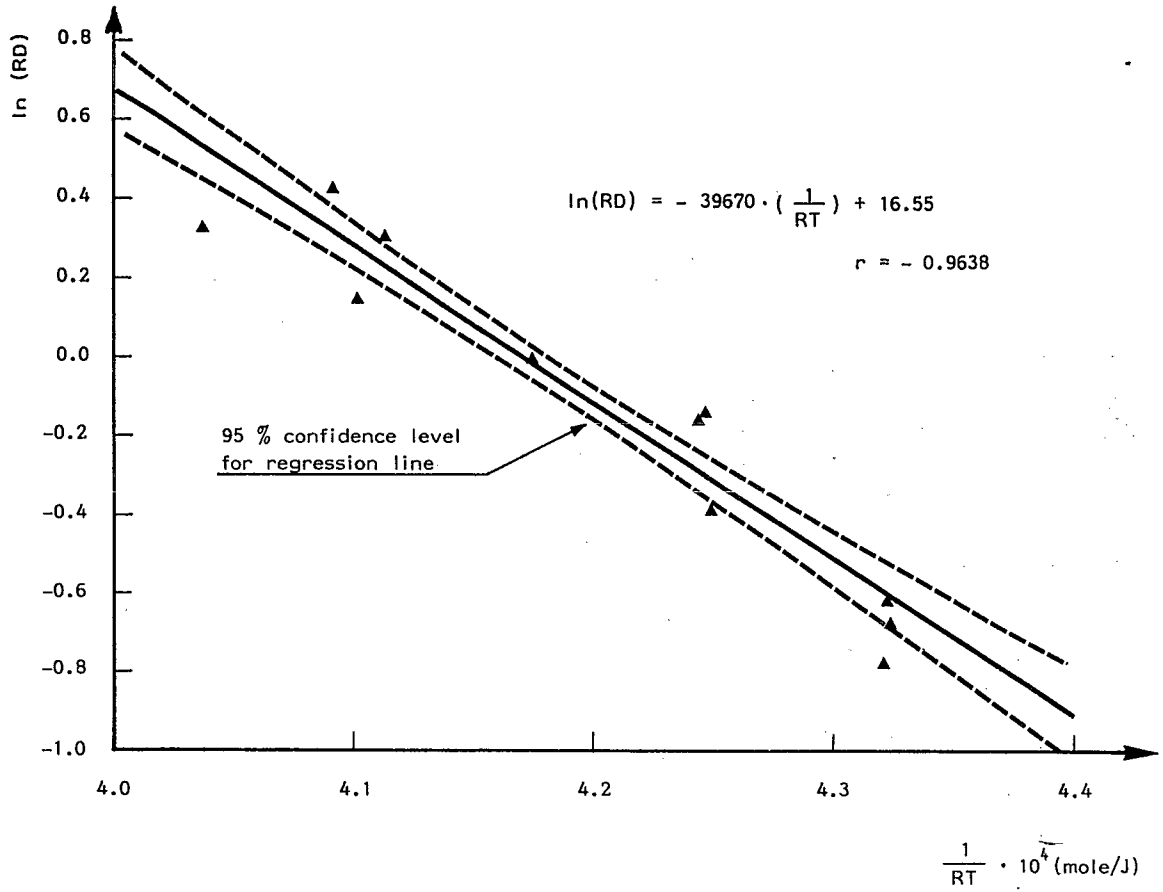

Figure 6. Arrhenius plot of the effect of temperature on denitrification.

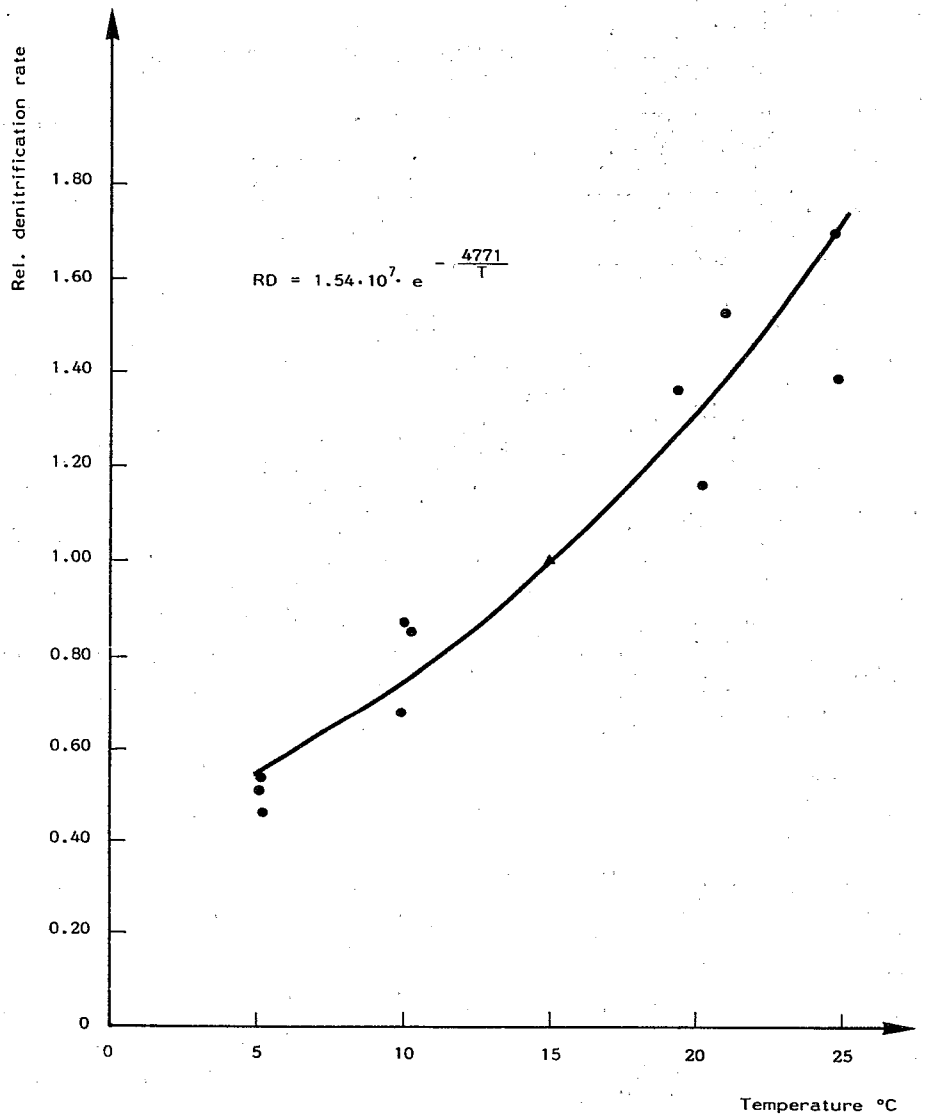

Figure 7. The effect of temperature on the rate of denitrification.

Table II. Temperature dependency.

Temp. range °C	Q_{10}	Θ
15 - 25	1.74	1.057
10 - 20	1.78	1.059
5 - 15	1.82	1.062

Murphy et.al (7) reports an activation energy of 69 300 J/mole for a submerged RBC, using methanol as carbon source. This gives a greater temperature dependency than we found in our study. Davies and Pretorius (8) observed a great drop in efficiency below 10°C, reporting a Q_{10}-value of 1.38 between 10 and 30°C and a Q_{10}-value of 13.06 between 5 and 10°C.

Our experiments show that denitrification with municipal sewage as carbon source can be achieved down to 5°C. For design, a temperature coefficient (Θ) of 1.06 should be appropriate.

pH dependency

RBC B was kept constant at pH 7, and the relative denitrification rate at this pH is put equal to 1.00. In RBC A the pH was increased in increments of 0.5 pH-units up to pH 10. Then the unit was acclimatized at pH 6.5 before a gradual decrease to pH 5.

Visual observations showed that the disc started to loose excessive amounts of sludge above pH 9 and below pH 6. When the inner walls were taken out for wash, they showed no sign of growth at pH 10 and pH 5. The denitrifying organisms were not killed at pH 5. pH 4.9 gave a relative denitrification rate of 0.11 (Figure 8). The following day we observed a relative rate of 0.30 at pH 5.2, which is a pretty good recovery.

This study demonstrates that the optimum pH for denitrification lies between 7 and 8.5, which is in agreement with the results reported by Davies and Pretorius (8).

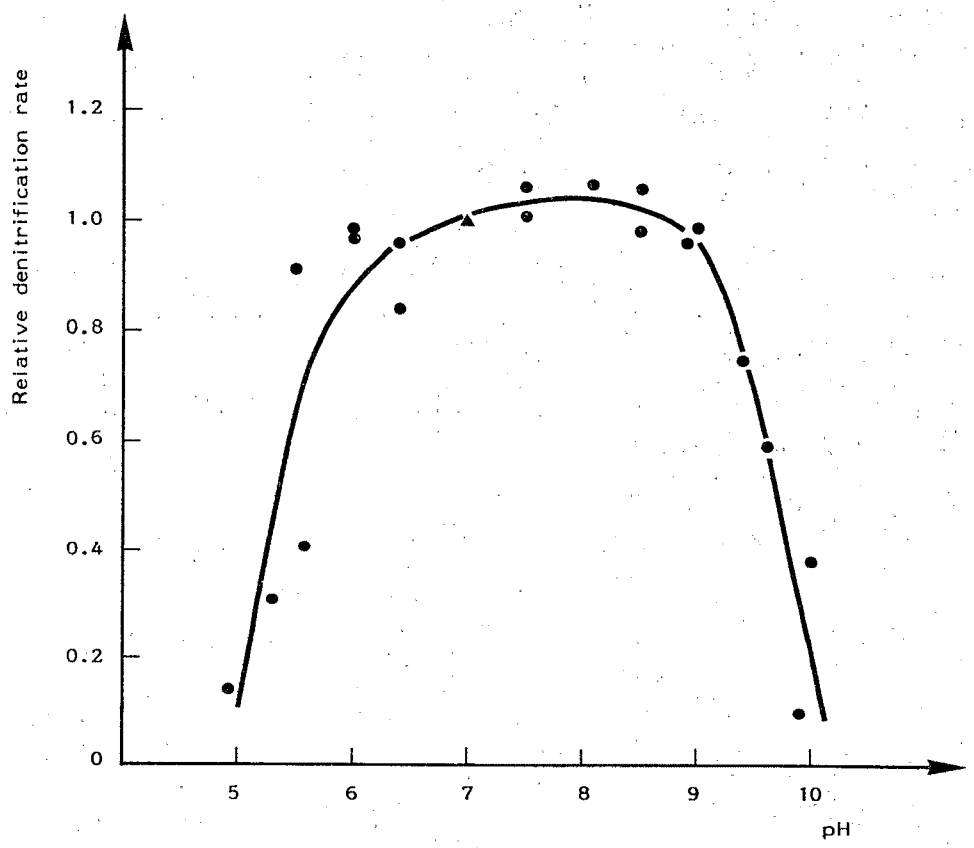

Figure 8. The effect of pH on the rate of denitrification.

Energy consumption

The energy needed for denitrification can be expressed by the substrate consumption ratio, defined as mg organic matter consumed/mg NO_x-N removed (9). Figure 9 shows a plot of NO_x-N removed versus $SBOD_5$ consumed, giving a substrate consumption ratio of 2.4 mg $SBOD_5$ consumed/mg NO_x-N removed. Using the regression equations in TableIII,the substrate consumption ratio in our study can also be expressed as 4.6 mg SCOD/mg NO_x-N or 1.6 mg DOC/mg NO_x-N.

Table III.Correlation between different organic parameters.

Regression equation	Number of observations	Corr. coeff.
$SBOD_5$ = 0.523 SCOD-11.4	138	0.9611
$SBOD_5$ = 1.524 DOC-12.9	133	0.9370

Narkis, Rebhun and Sheindorf (10) have published results from suspended culture experiments where different carbon sources (methanol, sodium acetate and chemically treated raw sewage) were used. They concluded that by expressing the organic matter as $SBOD_5$, a critical value of 2.3 mg $SBOD_5$/mg NO_x-N existed when 100% denitrification was to be reached regardless of what carbon source was used.

Monteith et.al (9) have investigated different carbon sources. They found substrate consumption ratios between 0.7 and 2.6 mg DOC/mg NO_x-N removed. Methanol had average values of 5.41 mg SCOD/mg NO_x-N and 1.17 mg DOC/mg NO_x-N. The substrate consumption ratios are influenced by the presence of dissolved oxygen and the carbon requirements for cell synthesis. A carbon source with high substrate consumption ratio, would tend to generate larger volumes of sludge, which is undesirable.

The municipal sewage used in this study has a substrate consumption ratio well inside the range found for other carbon sources (9,10). When a resirculation system (Figure 1) is used, influent dissolved oxygen concentration may increase, giving a slightly higher substrate consumption ratio. Using a ratio of 2.4 mg $SBOD_5$/mg NO_x-N removed will therefore be a

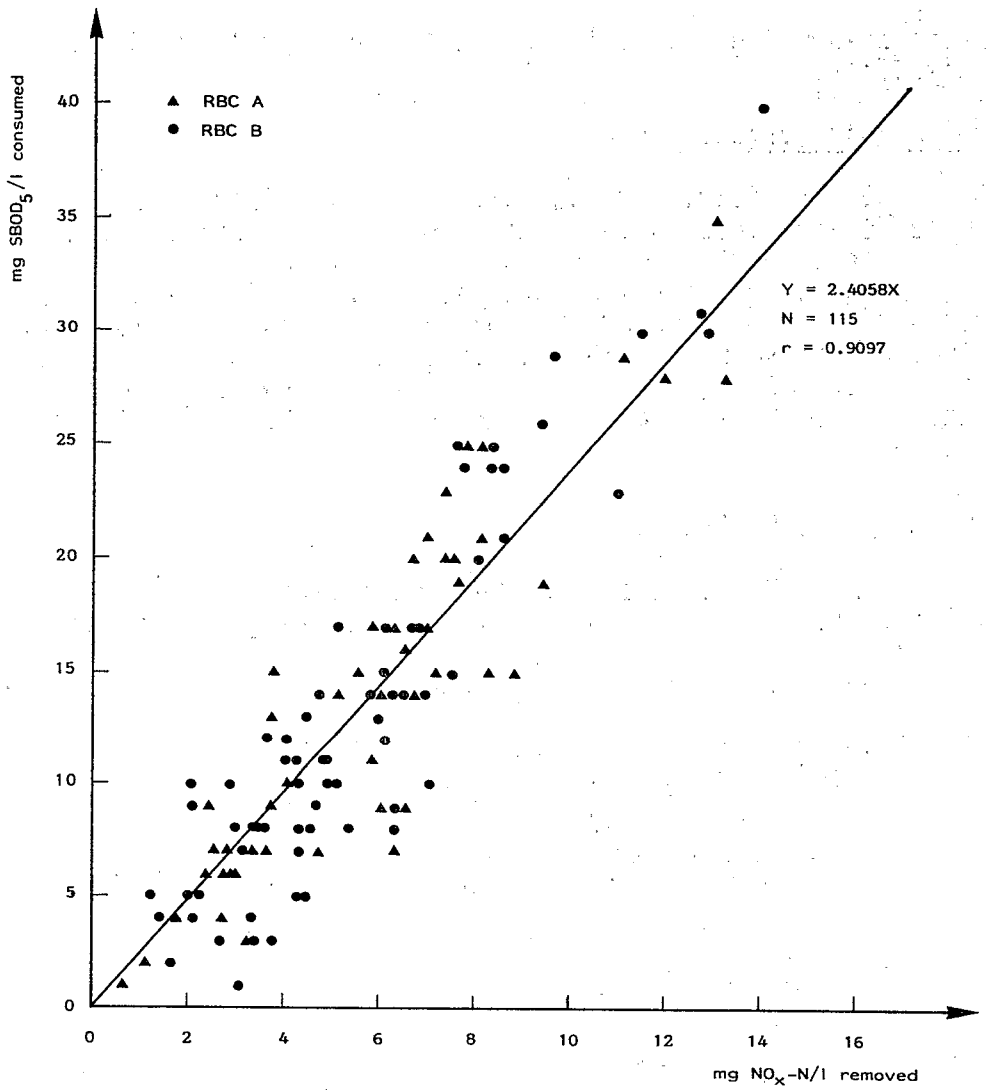

Figure 9. Energy consumption for denitrification.

safe assumption when calculating the organic load on the nitrifier.

Alkalinity production

Denitrification produces alkalinity. Theoretically 1 mole NO_3^--N removed gives 1 mole OH^-. This corresponds to an increase of 0.0714 meq/mg NO_3^--N removed.

Figure 10 shows the alkalinity production. It is found to be 0.0713 meq/mg NO_x-N removed, which is very close to the theoretical value. Using a RBC pilot-plant with methanol as carbon source, Smith and Khettry (11) observed a gain in alkalinity of 0.074 meq/mg NO_3-N removed (3.7 mg $CaCO_3$/mg NO_3-N).

SUMMARY AND CONCLUSIONS

Denitrification studies have been performed in two submerged bio-disc units, using presettled sewage as carbon source.

The conclusions are as follows:

1. The denitrification rate at 15°C and pH 7 could be described by the equation:

$$R_{NO_x\text{-}N} = 12.72 \left(\frac{L_{NO_x\text{-}N}}{335.3 + L_{NO_x\text{-}N}} \right) \cdot S_{O\ SBOD_5}$$, adding the limit

$$R_{NO_x\text{-}N} \leq L_{NO_x\text{-}N}$$

where

$R_{NO_x\text{-}N}$ = denitrification rate, mg/m²·h

$L_{NO_x\text{-}N}$ = NO_x-N load, mg/m²·h

$S_{O\ SBOD_5}$ = influent $SBOD_5$ concentration, mg/ℓ

2. For the system shown in Figure 1, denitrification can be regarded a 0.order reaction with respect to NO_x-N concentration, and a 1.order reaction with respect to $SBOD_5$ concentration.

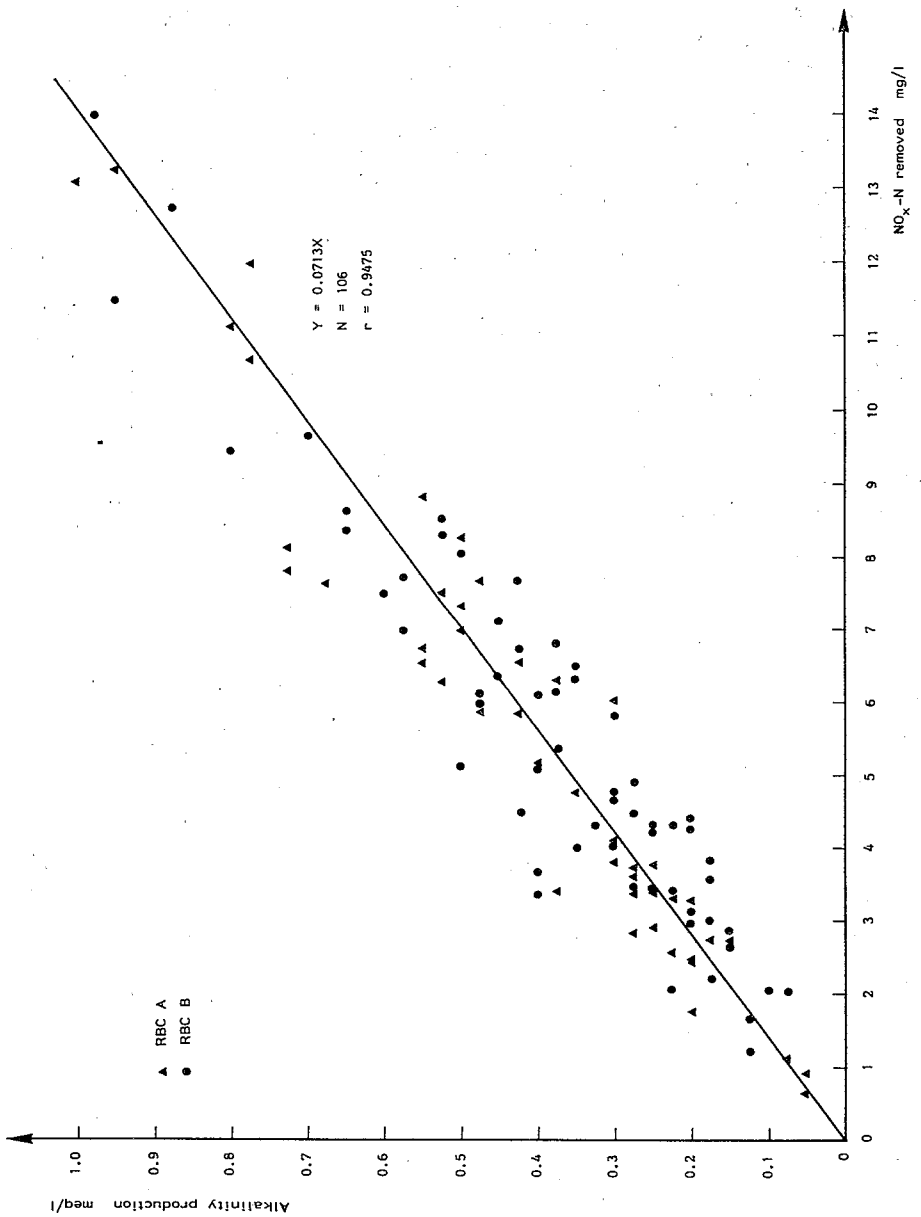

Figure 10. Alkalinity production.

3. Temperature dependency was modelled using an Arrhenius equation. This gave an activation energy of 39670 J/mole, corresponding to a temperature coefficient (Θ) of about 1.06.

4. Optimum pH lies between 7 and 8.5.

5. The energy consumption for denitrification has been found to be 2.4 mg $SBOD_5$/mg NO_x-N removed.

6. The alkalinity production has been found to be 0.0713 meq/mg NO_x-N.

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LIST OF SYMBOLS

Θ = temperature coefficient
$\hat{\mu}$ = maximum specific growth rate
d = thickness of the active biological layer
k = constant
r = correlation coefficient
A = area of biological film
A = frequency factor in the Arrhenius equation
C = constant
DOC = dissolved organic carbon
E = activation energy
K_S = saturation coefficient
$L_{NO_x\text{-}N}$ = NO_x-N load
NO_2-N = nitrite nitrogen
NO_3-N = nitrate nitrogen
NO_x-N = Σ NO_2-N + NO_3-N
Q = hydraulic flow rate
R = gas constant (8.314 J/mole·^{0}K)
$R_{NO_x\text{-}N}$ = denitrification rate
R_T = denitrification rate at temperature T
RD = relative denitrification rate
S_0 = influent substrate concentration
S_1 = effluent substrate concentration
$SBOD_5$ = soluble 5-day biochemical oxygen demand
SCOD = soluble chemical oxygen demand (Cr)
T = absolute temperature
X = concentration of organisms in the biological film
Y = yield

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OPERATION OF A RETAINED BIOMASS NITRIFICATION SYSTEM FOR TREATING AQUACULTURE WATER FOR REUSE

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ABSTRACT

A series of experimental trials were conducted in which a variety of polyurethane materials of differing pore size were evaluated as a media for nitrifying filters used in treating water in a trout hatchery. These filters were found to be capable of ammonia oxidation rates ranging from 80-180 mg-N/day/liter of filter volumes. These rates represent an order of magnitude increase over removal rates in conventional rock filters operating at influent ammonia levels at or below 0.5 mg/liter. These rates were, however, 50 to 100% lower than the rates observed for a similar filter design operating under laboratory conditions. The difference was attributed to increased heterotropic fouling experienced during the field operations. The optimum design for the field unit was found to be an initial stage of open pore media to accept the heterotropic loading followed by a second stage of fine pore material to allow for complete nitrification. Such filters may be operated at detention times less than four minutes achieving well over 90% ammonia removal.

INTRODUCTION

In the majority of cases in which low levels of ammonia or nitrite must be removed from water supply systems, one of two techniques have been utilized, either biological nitrification or breakpoint chlorination. Gauntlet (1) compared biological nitrification to breakpoint chlorination as a means of treating water for potable supply. He suggests that the disadvantages of relatively long contact time required for chlorination, compounded by interferences from organic compounds and potential production of dichloramines, trihalomethanes and other organochlorine compounds, makes nitrification a more desirable means of low level ammonia removal from potable water supply. Short (2) suggests that the fluidized bed nitrification process is cheaper than breakpoint chlorination when the ammonia levels to be removed exceed 0.2 mg/l.

Brune and Gunther (3) suggest that biological nitrification could, in fact, be used to economically remove low levels of ammonia from recirculating aquatic animal culture facilities. Studies conducted by Gunther et. al (4) indicate that ammonia levels ranging from 0.05 to 0.50 mg/l could be expected in such systems. The lower limit on these concentrations appear to favor breakpoint chlorination. Unfortunately, however, the dangers of chronic toxicity to fish from dichloramines or the possibility of acute toxicity from accidental chlorine overdose makes this an undesirable system for treating aquaculture reuse water.

In spite of the reduced rate of biological nitrification at these low ammonia levels, these sytems can be made to operate at high efficiency if the lower bacterial growth rate can be compensated for by carrying much higher levels of total biomass within the filter units, and at the same time, using a filter media which permits high water passage rates. Brune and Gunther (3) proposed the use of a "Retained Biomass Filter" for such systems. These filters consist of submerged, downflow columns of polyurethane cubes or sheets contained within ridge cells of plastic netting. When tested under laboratory conditions, such filters were capable of oxidizing ammonia at rates of 100-400 mg-N/day/liter of filter volume as compared to 10-30 mg-N/day/liter for conventional aquaculture filter designs.

The purpose of this study was to further examine the behavior of these retained biomass filters. In particular, this study was directed at examining the success of three such filter units under actual field conditions and to select the appropriate media pore size yielding optimum performance.

METHODS AND MATERIALS

This study consisted of three separate trial runs with a bank of filter units installed at the University of California - Davis trout hatchery previously described (4). Figure 1 illustrates the three filter units and the containment building. Figure 2 shows the placement of the filter units in relation to the trout culture and water storage tanks and rapid sand filters.

Each of the filter units consisted of a plexiglass tube 2 ft. deep by 1 ft. in diameter, sealed at each end with a removable plywood section. During the runs, the tubes were filled with varying types and configurations of polyurethane material.

Water exiting from the trout culture tanks would first pass through the solids settling tank (with approximately 2 minute detention time) and then be distributed to each of the three filter units. Flow rates to the individual filter units were controlled by 1/2 inch plastic valves and ranged from 3-7 liters/minutes giving filter detention times averaging 3-4 minutes. Each of the filters units was equipped with 7 sample ports at 2 inch intervals for sampling purposes. Total pressure drop across the filters was limited to 4 inches.

In the first experimental trial, each of the 3 units was filled with a different media type consisting of: 1) 1/2 inch square cubes cut from a dense polyurethane material with pore size of 0.1 mm. The cubes were loosely packed in six two inch deep ridge cells fabricated from "conwed" plastic netting material, 2) 1/2 inch square cubes in a similar arrangement but cut from a lighter polyurethane material of approximately 0.15 mm pore size, 3) a similar arrangement except the light polyurethane material (0.15 mm) was arranged as uncut circular sheets 1 inch thick stacked 24 deep. These filter units were operated for approximately three months. Inlet ammonia levels ranged from 0.25 to 0.40 mg/l depending on fish loading and water flow rates.

In the second experimental run, three different pore sizes were evaluated. These were a 2.0 mm, 1.3 mm, and 0.5 mm pore size in an open mesh polyurethane material ("Scott" brand industrial foam obtained from Wilshire Foam Products, Inc., Carson, California). In all cases, the material was distributed in the columns as 12 inch uncut circular disks 1 inch thick with 2 disks per ridge cell with a total of 24 disks. The ridge plastic screen was designed to prevent compaction of the filter media during operation. These filters were also operated for a period of 3 months with inlet ammonia levels ranging from 0.25 to 0.50 mg/1.

In the last experimental run, the three filter units were operated with a mixture of pore sizes: 1) arrangement A consisted of 100% 2 mm pore size, 2) arrangement B consisted of one half 2 mm pore size with one half of the 0.5 mm pore size, 3) arrangement C consisted of one third 2 mm, one third 1.3 mm, and one third 0.5 mm pore sizes. In all cases, the filter media were arranged with largest pore sizes toward the top of the column. This last run lasted 2 months and ammonia concentrated ranged from 0.10 to 0.35 mg/1.

During each run, water samples were routinely taken from inlet, outlet, and across the columns. Samples were taken every 3 to 7 days depending on operating conditions. Filter performance was evaluated by monitoring ammonia removal rates across the columns. Ammonia concentrations were determined by the phenolhypochlorite method of Solorzano (5) with modification of Liddicoat et. al (6). Reagent grade hypochlorite solution was used to make up the oxidizating solution as outlined by Solorzano. Liddicoat et. al (6) suggested that more consistent reagent blanks could be obtained by substituting potassium ferrocyanide as the catalyst in place of sodium nitroprusside. This modification of the phenolhypochlorite method was used in the present study.

Water samples were assayed within one hour of sampling. Preliminary tests showed that they lose about 1% of their ammonia per hour when stored at 18°C. No correction factor was used, however, in the calculation of ammonia concentration. Samples were simply assayed as soon as possible. A calibration curve was made for determining the ammonia concentration. The standard deviation of a 25 x 10^{-6} molar (0.35 mg NH_3-N/1) ammonia sample was $\pm$ 9% using one centimeter round curvettes and a B & L Spec 20 spectrophotometer. Nitrite was measured colorimetrically by the method of Strickland and Parsons (7).

RESULTS AND DISCUSSION

Figure 3 and Table 1 give data previously presented by Brune and Gunther (3) obtained from a laboratory nitrifying filter. Figure 3 shows that these units using a 6 inch downflow filter with a 0.15 mm pore size media were capable of ammonia oxidation rates as high as 250 mg-N/day/liter, and that these removal rates could be correlated reasonably well to the influent average ammonia concentration, if separated into detention time groups. In contrast, typical field filters used in aquaculture operation (Kramer, Chin, & Mayo; Table 1) operated at rates an order of magnitude lower. In fact, the previous laboratory results represented removal rates more closely approaching rates observed by Haug & McCarty (9) in filters operating at high ammonia input levels.

In comparison, the removal rates observed for the first series of field units (Figure 4) in this study averaged only around 40 mg-N/day/liter with total ammonia removal of 60-70% across the entire length of the column. The reason for this lower removal was obvious from visual inspection of the filters. The retained bimass filter operates on the principle that essentially 100% of the accumulated bacterial biomass is retained within the filter media. When these bacterial levels become restrictive to the water flow rate, the filter bed is washed, thus the biomass levels in the filter are externally controlled, rather than depending on sloughing rate, as in the case of conventional rock filters. Although the low pore size material permitted high nitrification rates and high flow rates in the laboratory, this was not the case in the field operation. The smaller pore operating under field conditions rapidly became clogged from heterotrophic growth as a result of the low levels of dissolved organics (~ 10 mg/l) and finely divided particulates present in the trout water. As a result, frequent washing of the filter media was required, drastically reducing the total nitrifying biomass. The 0.5-1.3 mm pore was judged inappropriate for field use and the run was abandoned after 3 months operation. However, the data (Figure 3) does give a good indication of the length of time required for filter start-up. As can be seen, approximately two months were needed to establish a completely nitrifying filter. The filters would typically rise to a high rate of nitrifi-

cation as the biomass levels increased, thus requiring washing of the media and afterward, weekly washing of the media with removal rates stabilizing around 30-50 mg-N/day/liter.

Figure 5 illustrates the performance of the 3 larger pore size polyurethane media (see Figure 6 for a media comparison). These three filters were seen to come up to full performance a month earlier than the previous run, most likely, a result of high level of bacteria in the system tanks and piping. Also, these filters because of their more open nature, were able to perform at a higher biomass level giving removal rates around 150 mg-N/day/liter and total ammonia reducations of 70-99% across the filter depth. The finer pore filter media showed greater fluctuations in removal rates since it tended to develop higher biomass levels followed by a need for more thorough cleaning. Figure 7 shows that removal rates could again be correlated reasonably well with influent ammonia levels.

When fully loaded at influent ammonia levels of 0.40 to 0.50 mg/l, these filter media were able to perform at 140-180 mg-N/day/liter. This rate is again roughly an order of magnitude higher than previous fixed bed nitrifying filters used at these low levels. However, these rates are only 1/2 of the rates observed in the laboratory. The primary difference between the laboratory and field performance is again the fouling of the filter due to the added heterotrophic bacterial loading.

In comparing the 3 pore sizes, one can see (Figure 8) that the 0.5 mm media was much more effective in removing ammonia requiring only 1/5 of the filter depth to oxidize the majority of the inlet ammonia, while the 2 mm media required the full depth to achieve significant removal. Perhaps, just as important is the lower level of nitrite reduction in the 2 mm media (Figure 9). Although the output level of nitrite from the 2 mm media was low in relation to the total ammonia and nitrite oxidized (Figure 10), these levels are still reason for concern since low levels of nitrite also present a chronic toxicity problem to fish.

Finally, Figures 11, 12, and 13 show data taken from run three in which the media types were mixed. Filters B and C represent the most successful modification. As seen in Figure 11, removals ranged from 80 to 120 mg-N/day/liter. This removal efficiency is at the same level as the previous runs when consideration is given to the reduced influent ammonia levels experienced during these trials as a result of lower fish loading during this period (Figure 14). Figure 12 shows the same correlation between filter performance and average influent ammonia levels as previously demonstrated.

The important advantage of the combination of pore sizes in the filter were: 1) the filter could be maintained at high flow rates with only a weekly or biweekly washing of the filter media since the heterotrophic biomass was carried in the upper, more open pored layers and 2) a high level of nitrification including more complete nitrite removal could be maintained in the lower, smaller pored media.

SUMMARY AND CONCLUSIONS

As a result of a series of pilot tests conducted in a typical trout culturing facility, a more optimum design of a biological nitrifying filter for treatment of low level ammonia laden water has been achieved. The data obtained in this study indicates that a reasonably high rate of ammonia and nitrite oxidation (140-180 mg-N/day/liter) can be achieved with an approximately 3-4 minute detention time when using a filter design consisting of 6-10 inches of 2 mm open polyurethane followed by 6-10 inches of 0.5 mm polyurethane filter media. These filters act as complete biomass capture systems and require weekly to biweekly solids removal for optimum performance.

The initial larger pore section of this filter acts as a partial nitrifying and complete organic reduction unit, while the second smaller pore section allows for more complete nitrification. The two stage design allows for a sustained high water passage rate requiring a minimum of filter washing.

Additional work should be conducted to develop a low cost, automatic design for controlling the biomass washing of these filter media in large scale units. Secondly, a filter media material is needed that is more resistant to degradation. After one year of operation all polyurethane media used in this study had undergone severe degradation, so much so, that they would have needed replacement had the experiments continued for a longer period of time.

ACKNOWLEDGEMENTS

This research was made possible by funds provided by the University of California - Davis Aquaculture Program. At the time of this study D. E. Brune was Assistant Professor of Agricultural Engineering at UCD. R. Piedrahita was a Research Assistant in Agricultural Engineering at UCD.

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	Inlet Ammonia Levels (Mg/l)	Loading Rate Mg-N/D/L	Removal Rate Mg-N/D/L	Detention Time (Min.)	Temp. °C
Kramer, Chin, & Mayo (8)					
BTF	0.45	37	11	17.6	Variable
UF	0.84	85	17	14.2	Variable
AUF	0.80	73	25	15.9	Variable
DDF	1.77	90	30	28.2	Variable
SPDF	0.82	46	10	25.8	Variable
Haug & McCarty (9)					
Run 1	8.0	1536	1045	7.5	25°
Run 2	8.0	768	595	15.0	25°
Run 3	7.7	370	350	30.0	25°
Forster					
Run 1	1.0	50	49	28.8	26
Run 2	1.0	100	92	14.0	26
Run 3	1.0	200	170	7.2	26
Brune & Gunther					
Run 1	0.5	100	99	11.9	Variable
Run 2	0.5	240	220	5.1	Variable
Run 3	0.5	500	450	3.0	Variable

Table 1: Comparison of Various Filter Performances Under Differing Operating Conditions. (From (3)).

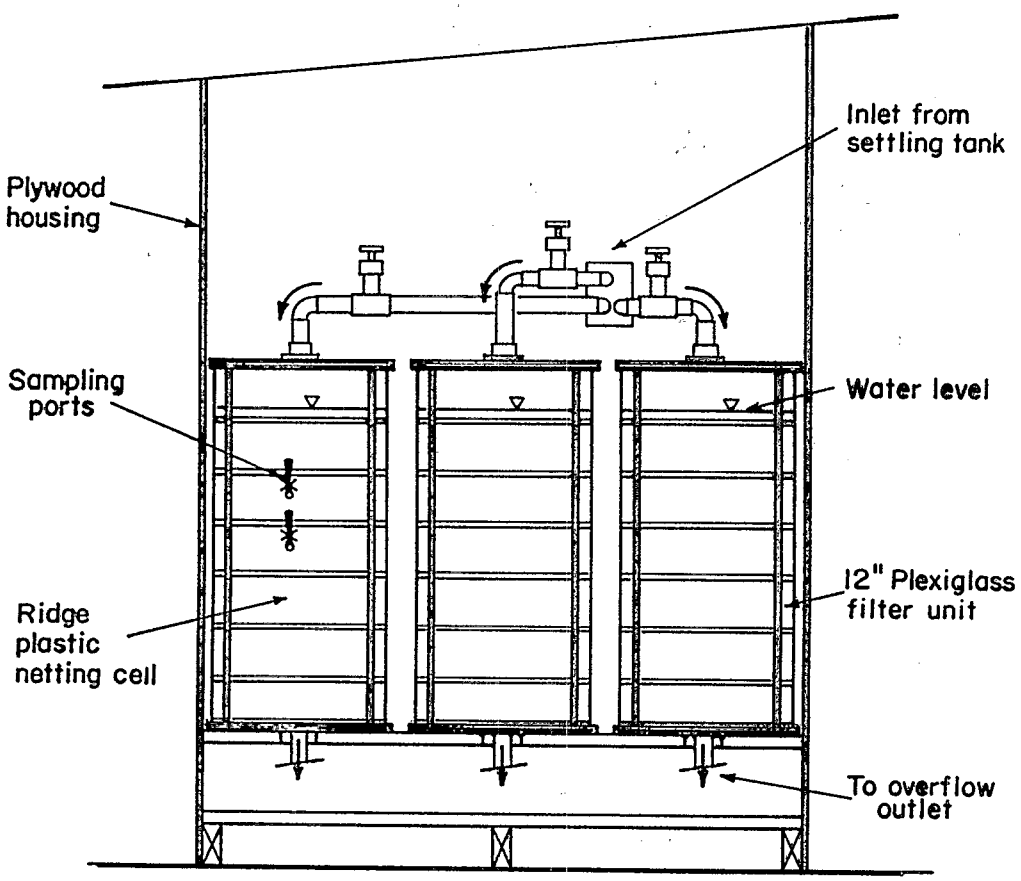

Figure 1. Experimental filter units.

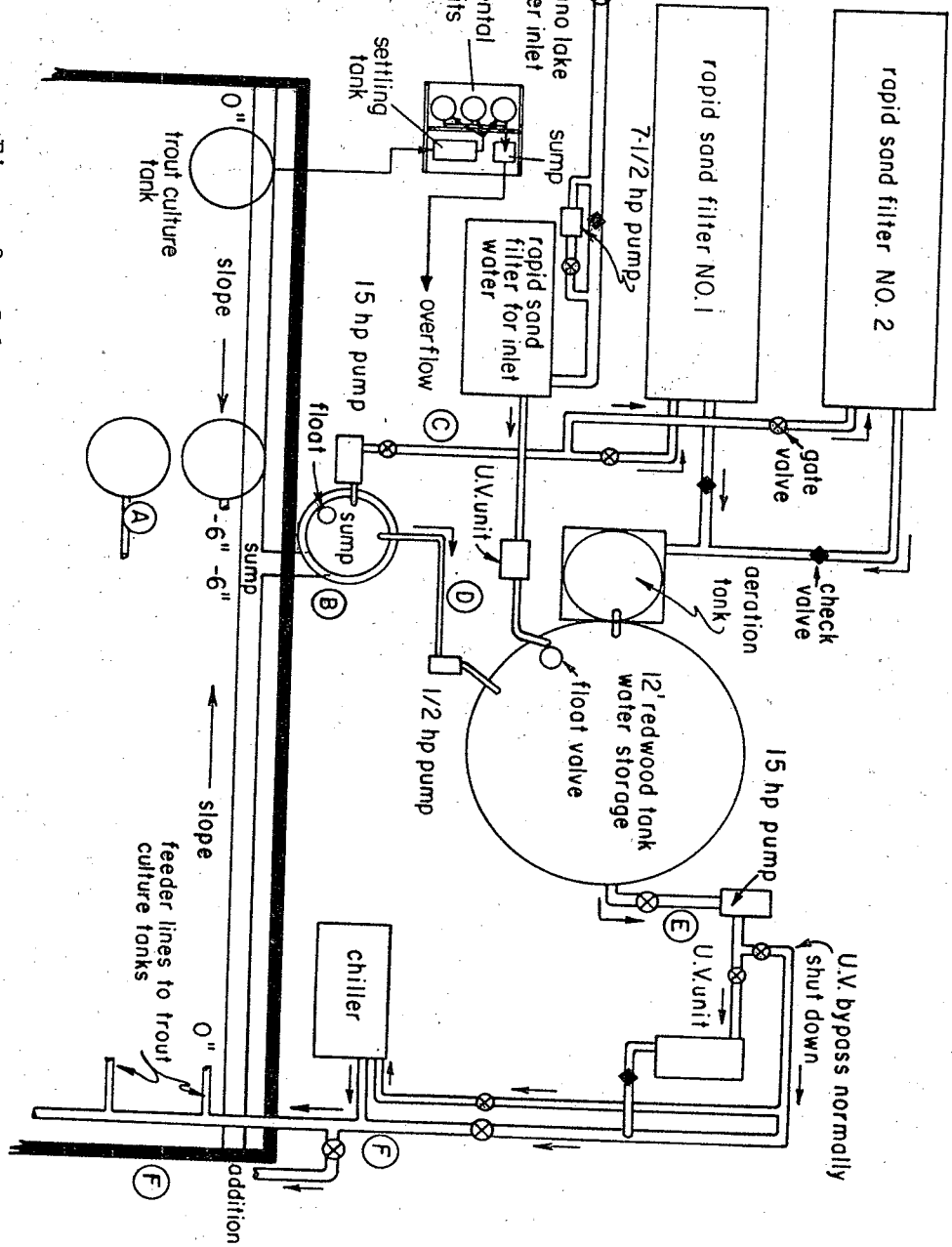

Figure 2. Schematic of University of California - Davis trout hatchery.

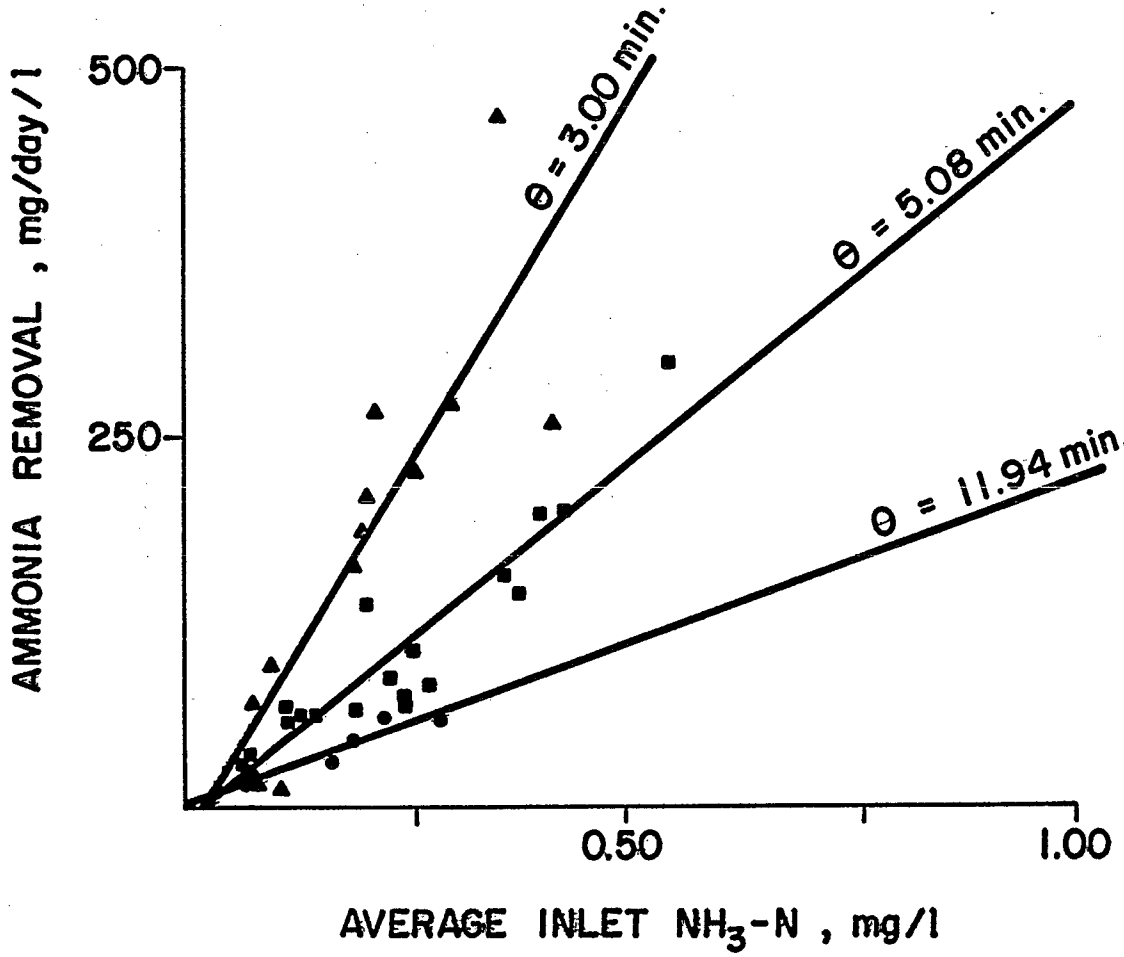

Figure 3. Average inlet ammonia levels vs. ammonia removal for laboratory filters (From Brune and Gunther (3)).

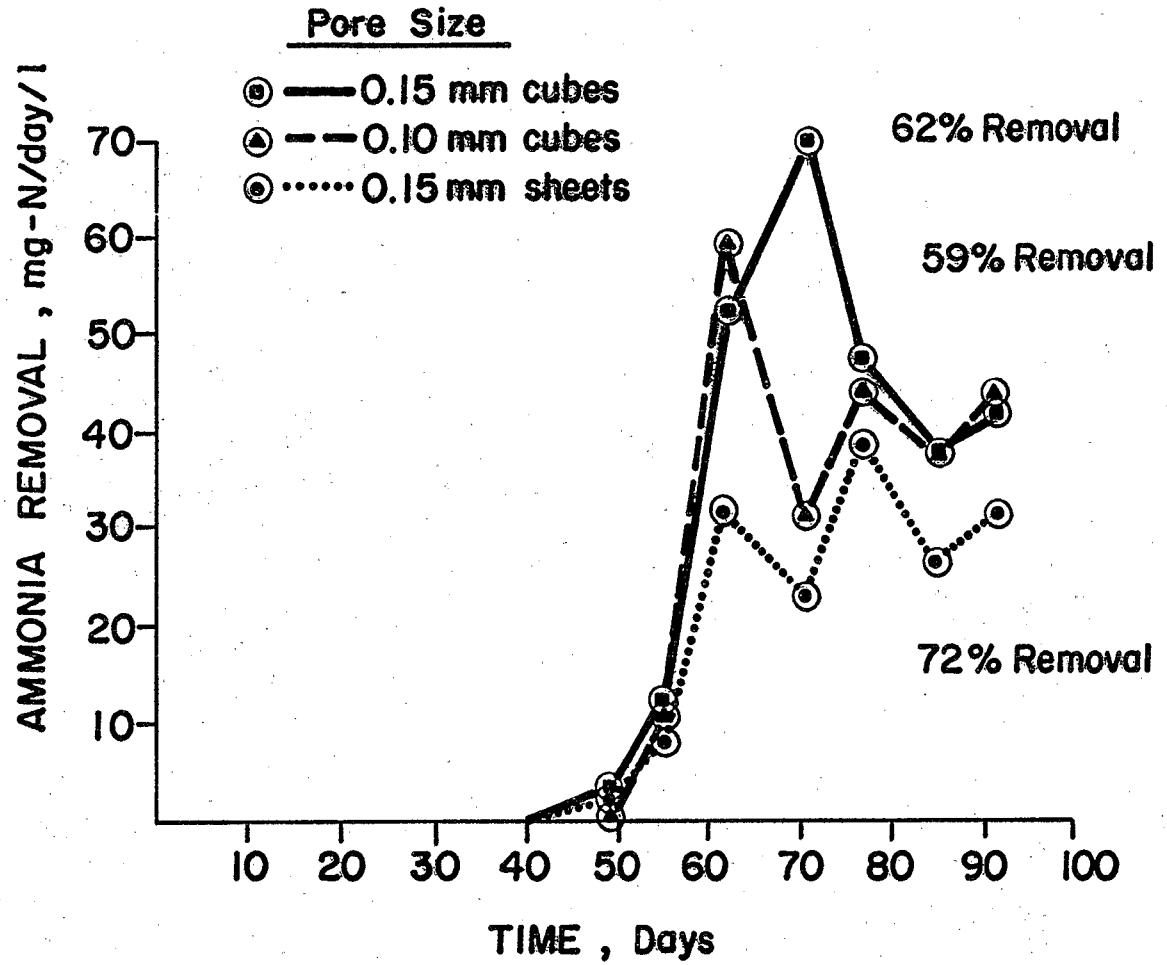

Figure 4. Ammonia removal in fine pore filters.

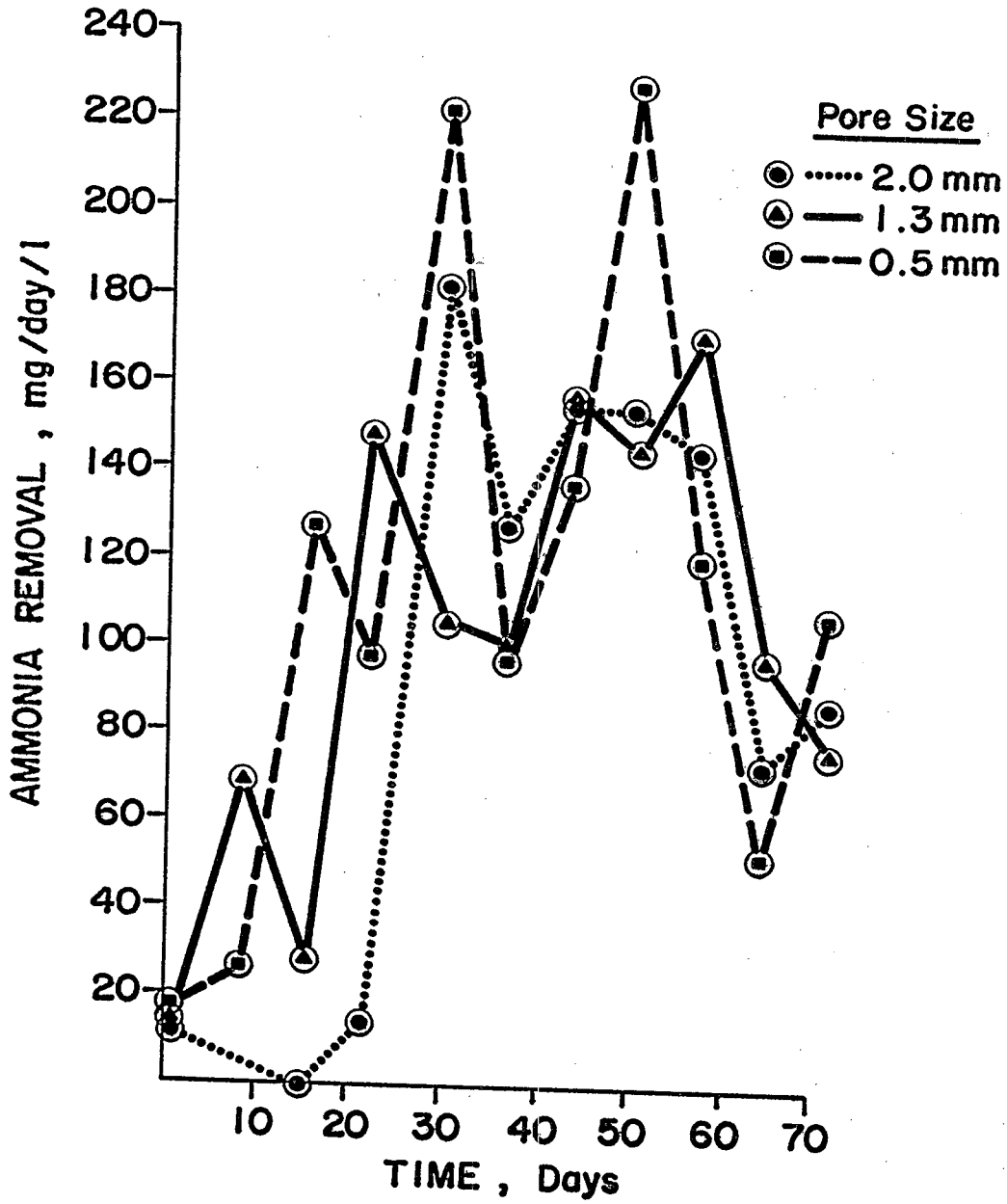

Figure 5. Ammonia removal in large pore filters.

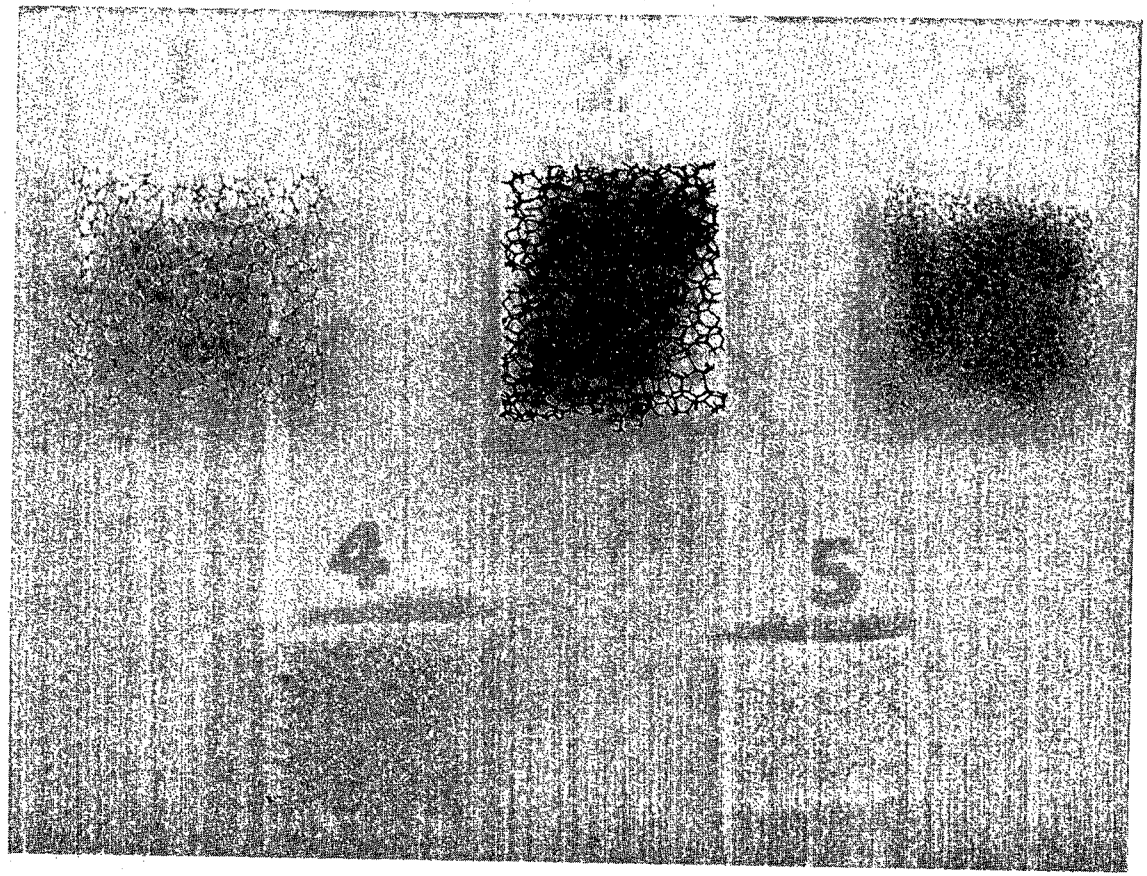

Figure 6. Various polyurethane materials used for filter media (1.5 x actual size).

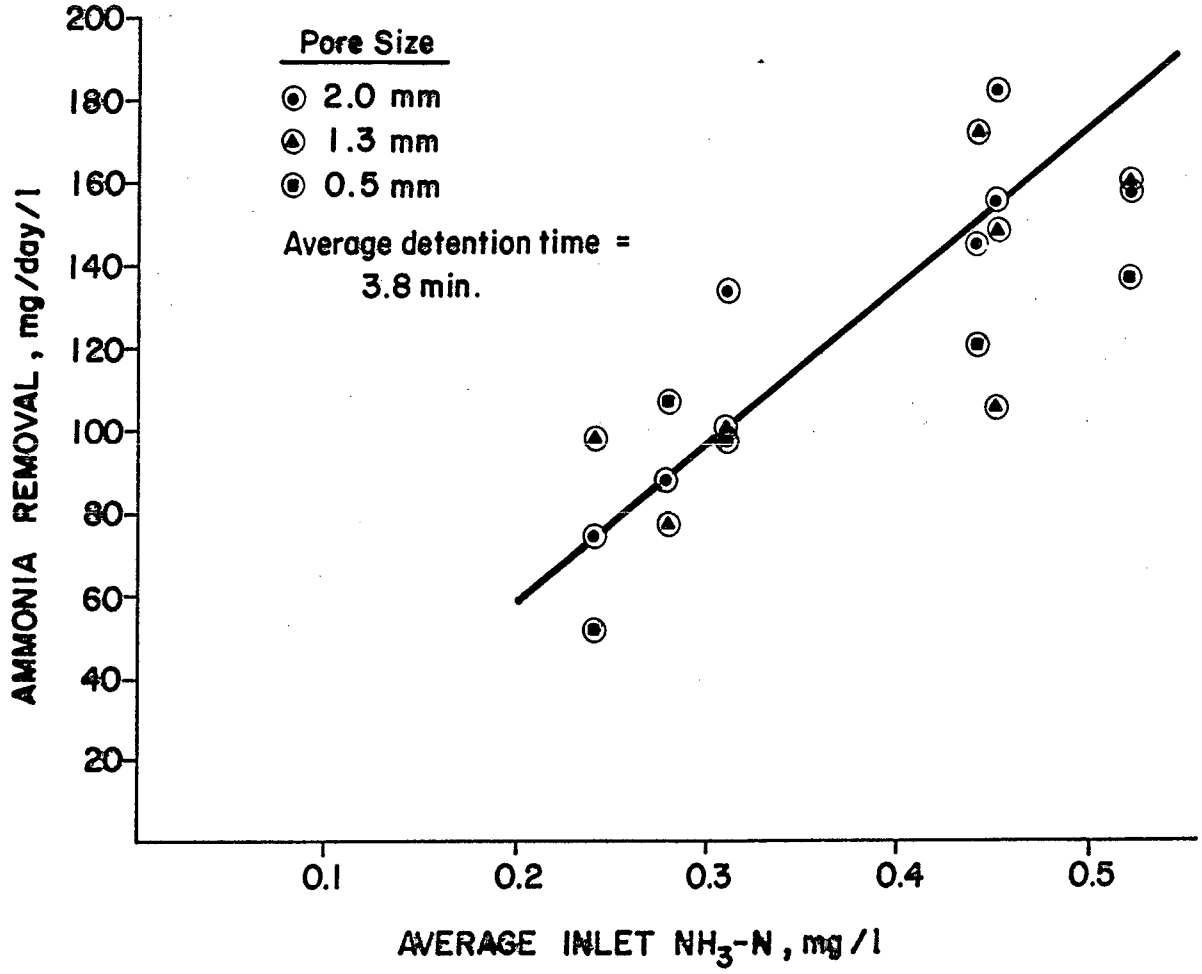

Figure 7. Relationship between ammonia removed and inlet concentration in large pore filters.

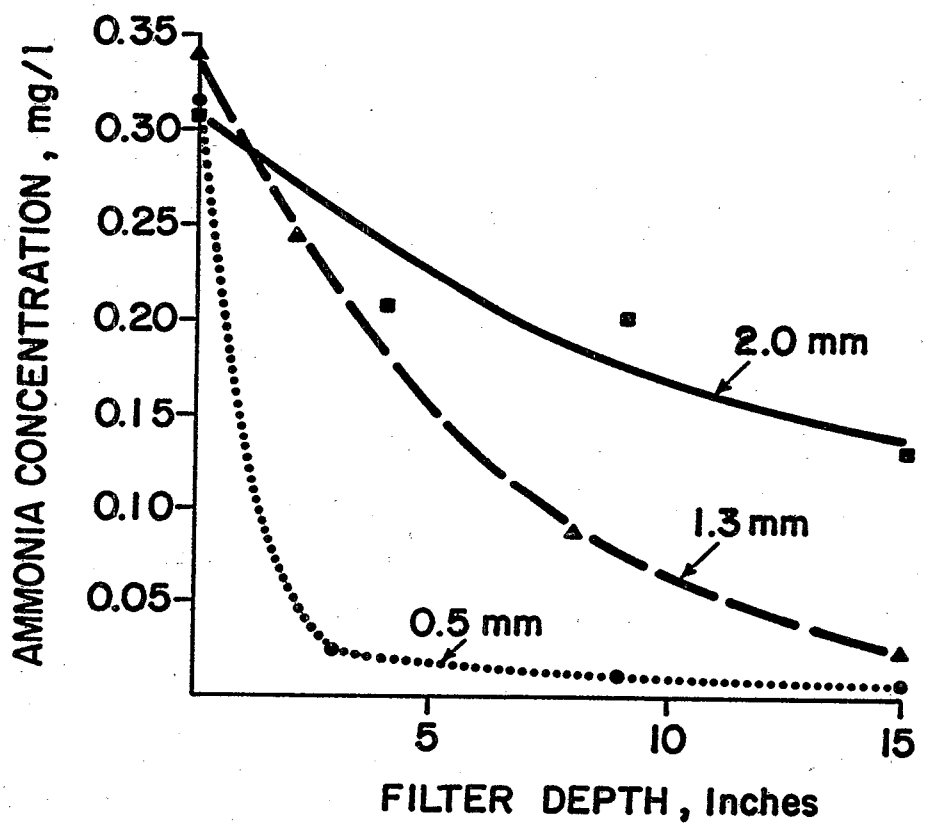

Figure 8. Drop in ammonia concentration across three different pore size filters.

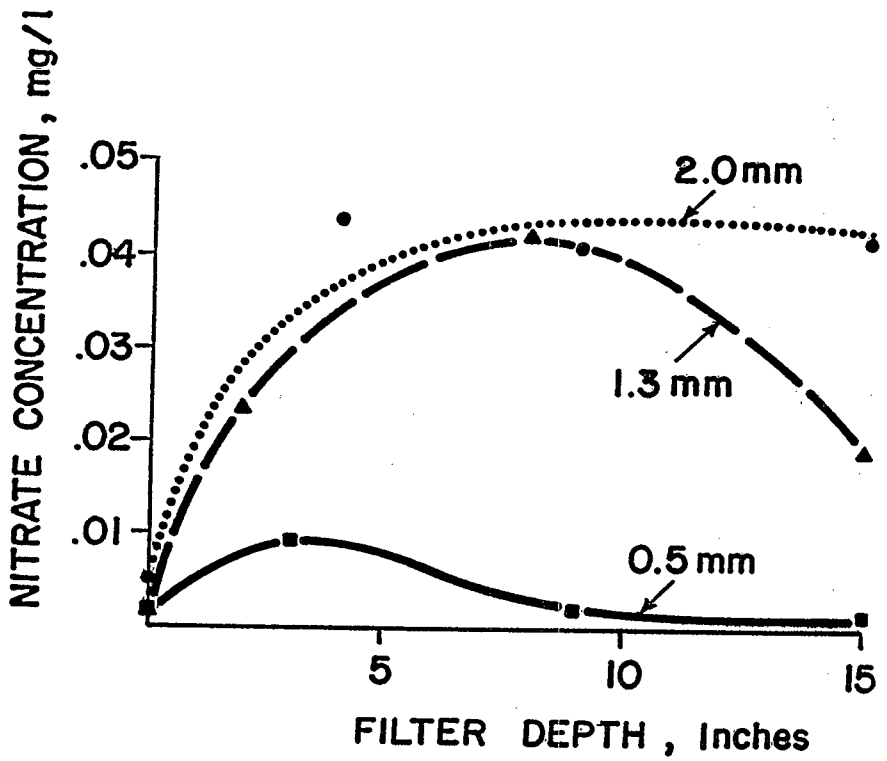

Figure 9. Levels of nitrite across three different pore size filters.

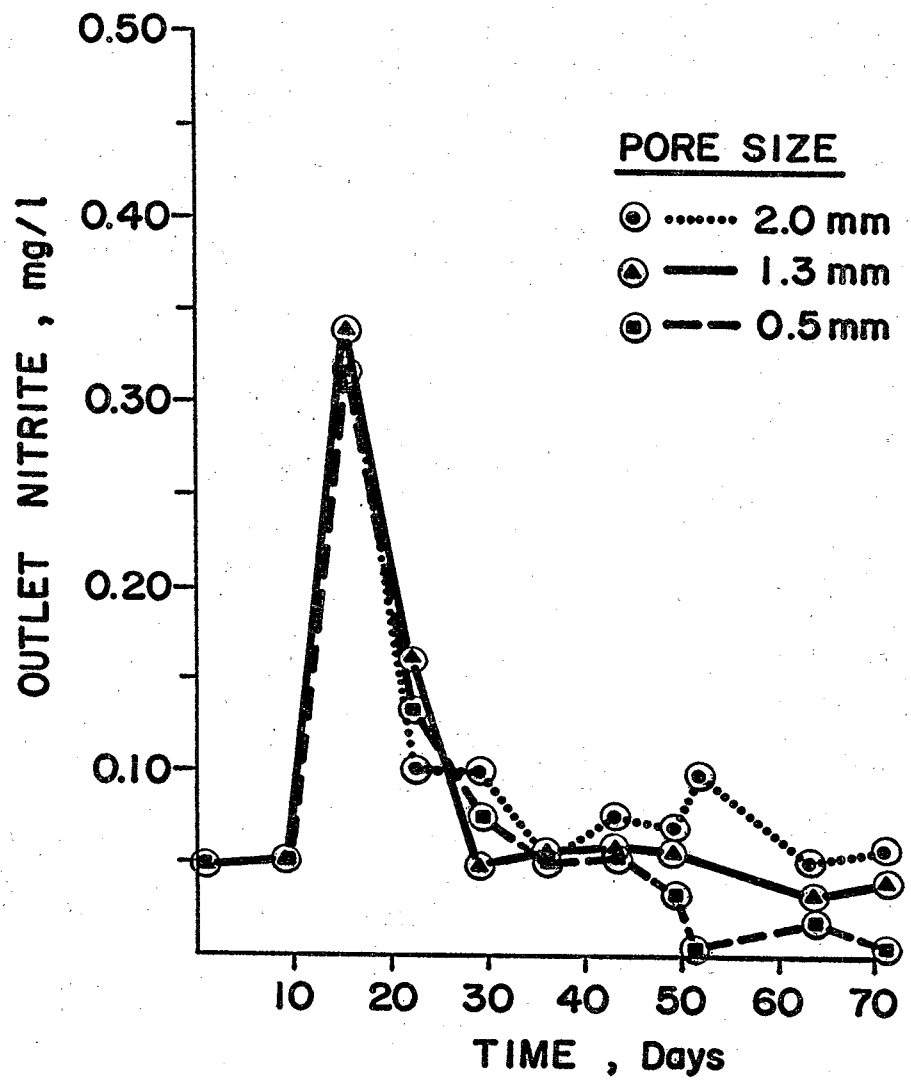

Figure 10. Levels of outlet nitrite from large pore filters.

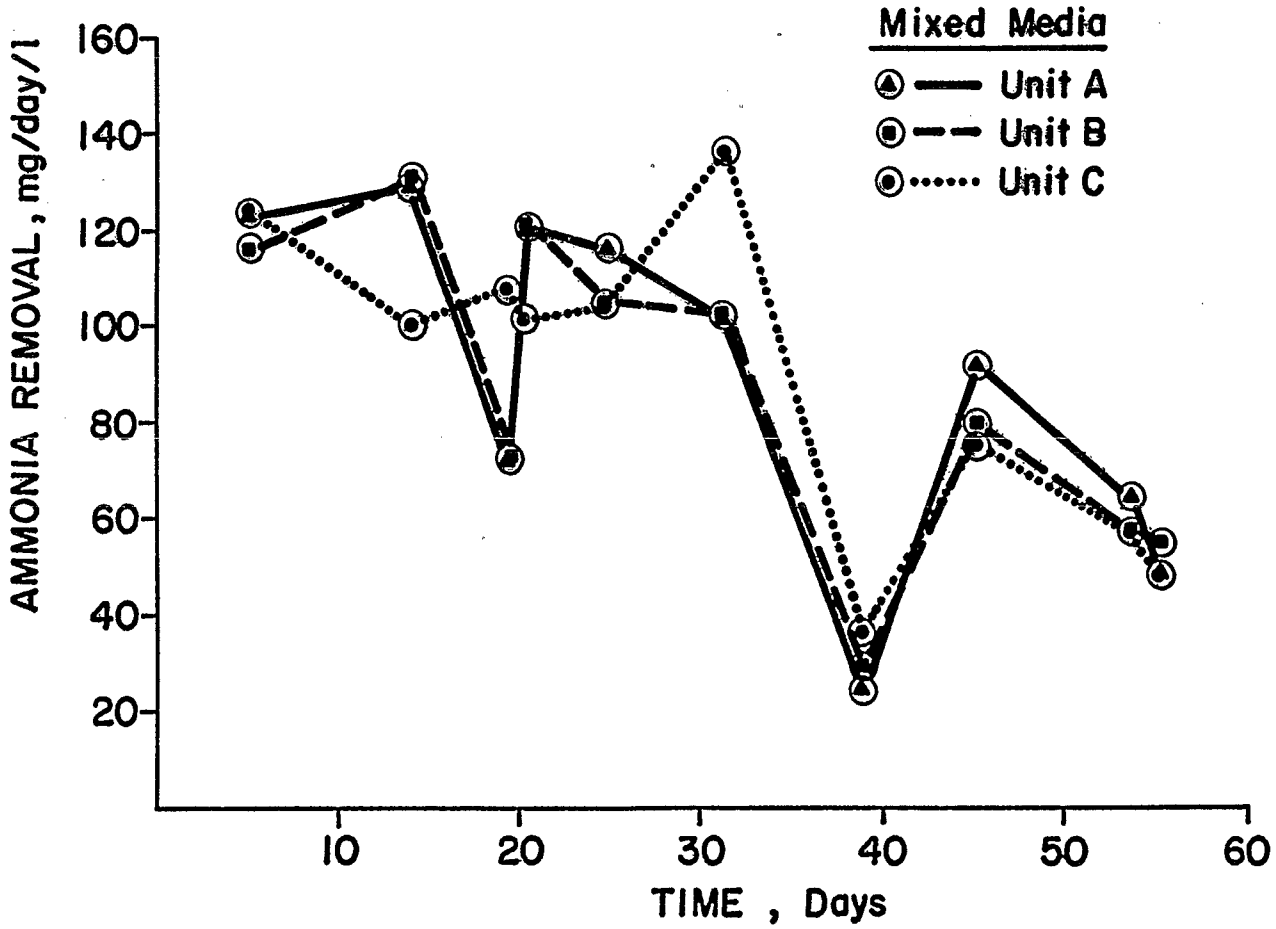

Figure 11. Ammonia removal in mixed media filters.

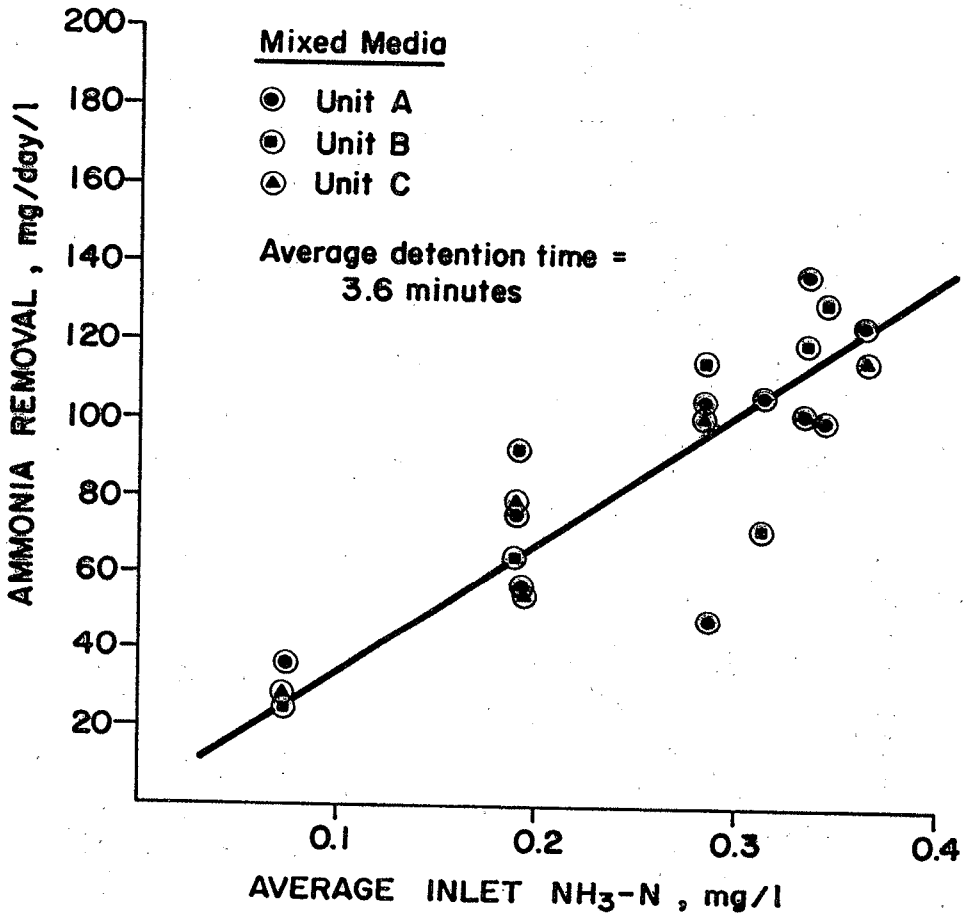

Figure 12. Relationship between ammonia removal and inlet levels in mixed media filters.

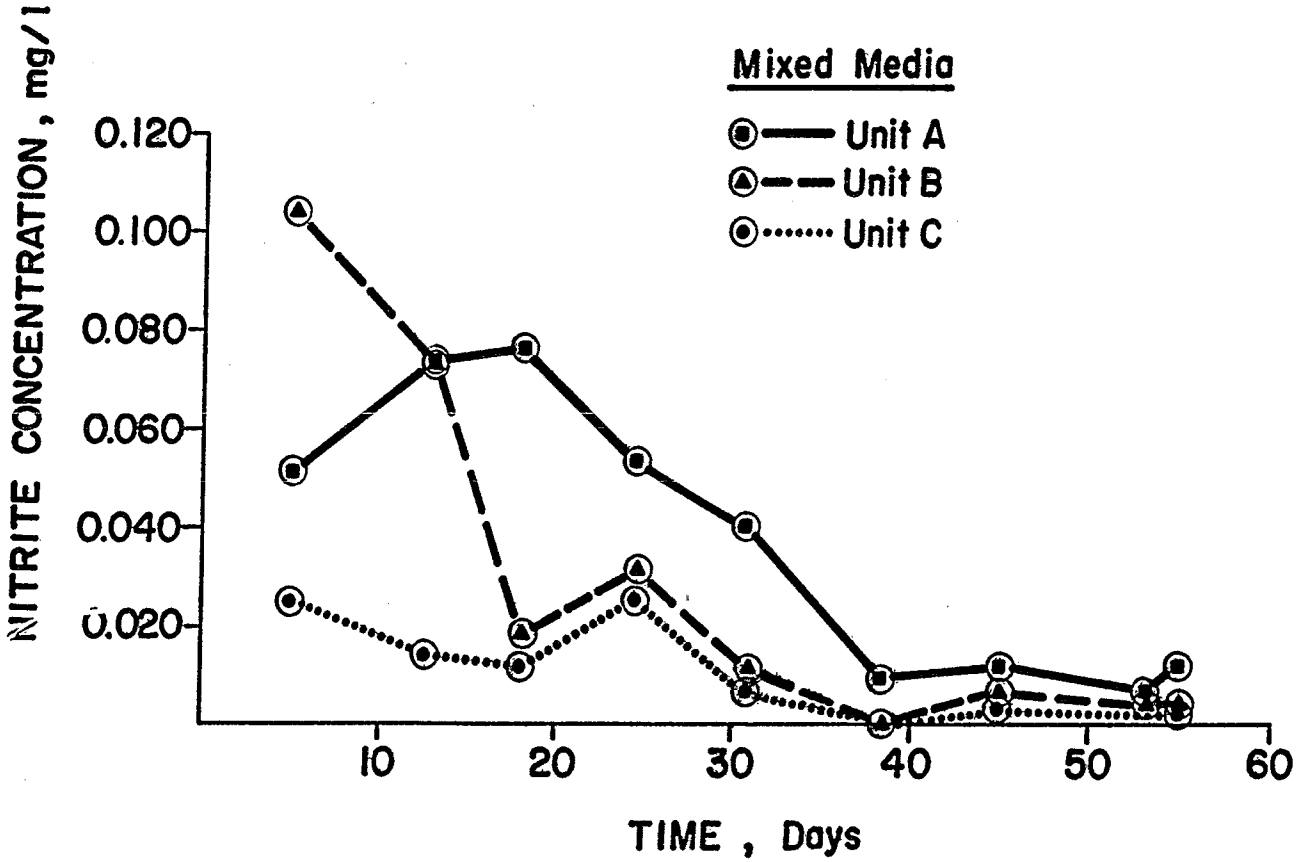

Figure 13. Outlet nitritie concentration from mixed media filters.

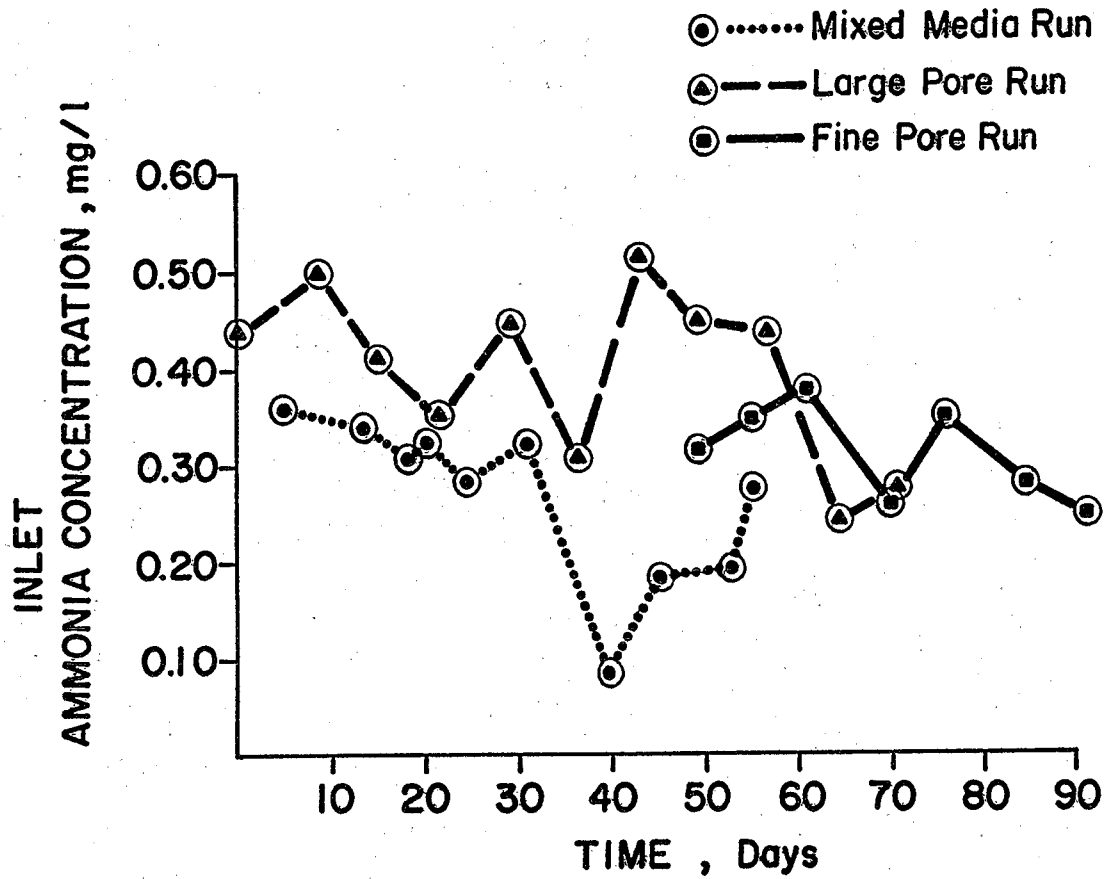

Figure 14. Inlet ammonia levels during the three trial runs.

Nitrified Secondary Treatment Effluent by Plastic-Media Trickling Filter

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INTRODUCTION

Ammonia nitrogen plays a vital role in the synthesis of microorganisms in the secondary treatment of wastewater. Generally, there is abundant ammonia nitrogen present in municipal effluent streams. Through the process of nitrification, this excess nitrogen in the wastewater treatment plant effluent consumes the dissolved oxygen in the receiving water. This, along with the ammonia toxicity, creates unhealthy conditions for aquatic life.

Nitrification is a two-step biological process in which ammonia nitrogen is oxidized to nitrite and further to nitrate as shown below:

$$NH_3 \xrightarrow[\text{Nitrosomonas}]{\text{Step one}} NO_2 \xrightarrow[\text{Nitrobacter}]{\text{Step two}} NO_3$$

Nitrosomonas and nitrobacter are commonly the most responsible for either of the two oxidation steps. These organisms are classified as autotrophic or chemolithotrophic because they

obtain carbon source from dissolved carbonate and use oxidizable substrate as their source of energy for growth and metabolism. The two steps for the nitrification transformation can be written as follows:

$$NH_4^+ + 1.5\ O_2 + 2\ HCO_3^- \xrightarrow{\text{Step one}} 2\ NO_2^- + 2\ H_2CO_3 + H_2O \qquad \text{----------(1)}$$

$$NO_2^- + 0.5\ O_2 \xrightarrow{\text{Step two}} 2\ NO_3 \qquad \text{----------(2)}$$

In addition, the overall reaction is

$$NH_4^+ + 2\ O_2 + 2\ HCO_3^- \longrightarrow NO_3^- + 2\ H_2CO_3 + H_2O \qquad \text{----------(3)}$$

Biological nitrification of a secondary effluent can be accomplished by both suspended- and fixed-growth processes (1-3). However, due to many advantages associated with the latter process, more studies have been made recently in the fixed-growth systems (4-7). Also, the use of light weight synthetic media allows the fixed-film biological filter to be constructed much deep thus minimizing space requirement and maximizing loading capacities. The synthetic media has an increased specific surface area for greater biomass attachement that results in a lower solids production in the plant effluent.

The main objective of this study is to investigate the feasibility of using plastic-media trickling filter for the removal of nitrogen from the effluent of activated sludge wastewater treatment plant. The efficiency of the trickling filter plant was evaluated under different nitrogen loading rates. And the relationship between alkalinity destruction and nitrogen removal was also studied. The settling charact-

eristics of the nitrified sludge was determined by sludge volume index (SVI).

EXPERIEMNTAL PROCEDURE

Two laboratory-scale trickling filter towers were constructed in the Environmental Engineering Research Center at the University of Pittsburgh. The filters were 6" squares and 8' long. A synthetic plastic media manufactured by Munters was employed for the study. The media model is Munters Biodek 19060, which has a surface area 140 m^2/m^3 or 44 ft^2/ft^3 and a void ratio of greater than 95%. Plant specifications and media structure are shown in Table 1.

The influent for twin towers, connected in series, was stored in a 250 gallons tank. The influent feed solution was taken from the effluent of an existing activated sludge pilot plant located in the same laboratory. The desired amounts of ammonia nitrogen (NH_4Cl) and sodium bicarbonate ($NaHCO_3$) were added to the solution and mixed in the feed tank.

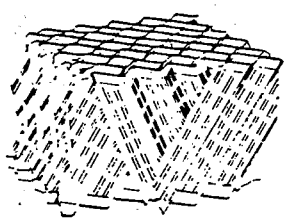
The wastewater was pumped to the plastic-media trickling filters by two variable speed pumps. The wastewater was first fed to the top of filter No.1 and the effluent was discharged to clarifier No. 1. And then a second pump taken the feed from the above mentioned clarifier, continuously pumped the wastewater up to the top of filter No. 2 where it traveled down and into clarifier No. 2.

Effluent samples were collected twice each week. Sampling ports were located 2 ft, 4 ft, 5.5 ft, and 7 ft from the top of filter media in each tower. Controlling parameters included NH_3-N, NO_2, NO_3, alkalinity, pH, dissolved oxygen (DO), suspended solids, BOD, and SVI. NH_3-N, NO_2, and NO_3 were measured by an Orion Digital Ionalyzer. Other parameters such as TSS and VSS, alkalinity, BOD were performed in accordance with "Standard Method" (8). The pH and DO were measured by Orion pH meter and YSI DO meter.

RESULTS AND DISCUSSION

1. Influent Wastewater Characteristics. Table 2 is a list of influent feed conditions proceeding into first filter. The range of flow rates tested varied from 45 to 115 gallons per day. The highest NH_3-N concentration was 109.4 mg/l while the lowest concentration was 43.6 mg/l. The influent nitrite and nitrate concentration ranged from 3.2 to 12.4 mg/l and the alkalinity fluctutated between 315.7 mg/l and 750 mg/l as $CaCO_3$. The pH range was small and it varied between 7.7 and 8.3. The dissolved

Table 1. Trickling Filter Plant Specifications

Parameters	Filters #1 and #2
(A). Reactor Dimension:	
Size ----------	6" x 6"
Height ----------	7'
Volume	
Gross ----------	2.0 ft^3
Media ----------	1.75 ft^3
(B). Filter Media:	
Type ----------	Munters Biodek I9060
Surface Area ----------	44 ft^2/ft^3 (140 m^2/m^3)
Void Ratio ----------	> 95%
(C). Media Configuration:	

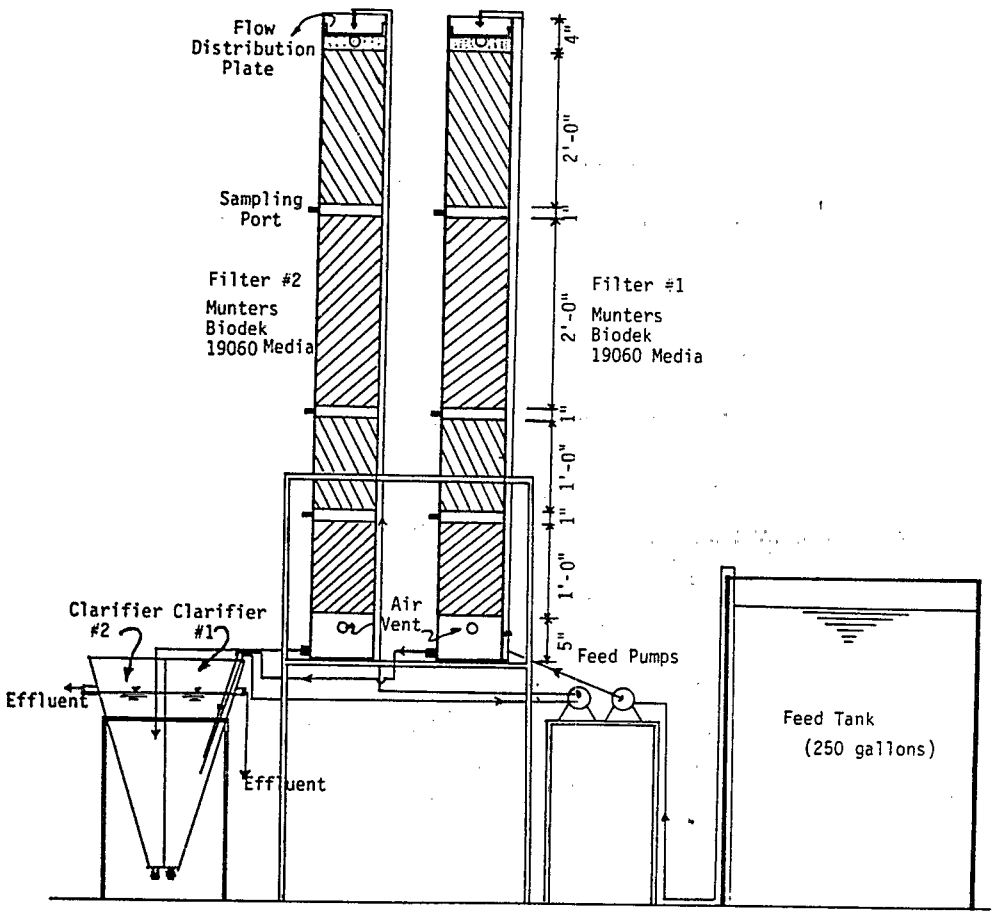


Figure 1. Trickling Filter Pilot Plant
For Nitrification Study

Table 2. Influent Feed Conditions*

Flow Rate	Influent Concentration, mg/l					
Gal./Day	NH_3-N	NO_2+NO_3	DO	Alkalinity as $CaCO_3$	pH	Temperature, °C
78.6	74.18	6.6	4.8	431.0	8.2	24.0
49.3	109.40	6.1	4.7	750.0	8.0	20.0
102.8	50.1	7.3	6.1	368.0	7.8	24.0
49.4	79.3	12.4	5.0	650.5	7.9	21.5
98.5	39.3	9.6	8.8	315.7	7.7	22.0
78.8	45.5	6.4	4.5	460.0	8.0	23.0
99.6	26.0	9.1	7.7	323.2	7.7	22.0
78.8	28.4	8.1	5.3	373.6	8.1	23.1
45.0	54.1	11.0	5.4	536.9	7.9	20.6
49.9	43.4	6.6	4.8	457.0	8.2	22.0
98.0	19.5	9.2	8.9	311.2	7.7	22.0
115.0	15.2	5.1	4.8	366.4	8.3	24.4
57.5	13.6	7.2	4.9	425.4	7.9	23.5
57.7	12.6	3.2	5.2	366.4	8.2	26.0

* BOD < 25 mg/l

oxygen content of the feed was within the level of 4.7 to 8.9 mg/l. The pilot plant was operated at room temperature, 20.6 to 26.0°C.

2. System Start-Up. At the beginning of the study, the trickling filters were operated separately at the flow rate of 42.6 gallons per day and 56.4 gallons per day. Figure 2 shows the NH_3-N and NO_2+ NO_3 concentrations at different filter depth with respect to days of operation for each filter.

From the figure it can be seen that it required 60 days for one filter and 67 days for the other filter to remove the NH_3-N down to approximately 1.6 mg/l at a filter depth of 6 feet. Also, from the same figure, it is apparent that the NH_3-N concentration decreases as the filter depth and start-up days increase. In addition, Figure 2 further shows that nitrification increases with filter depth and start-up time because there is an increase in NO_2 + NO_3 concentration.

As already pointed out, the start-up time for obtaining a steady-state operational condition in trickling filter nitrification process was long with no seed organisms used. In order to shorten the start-up time, flow recirculation through the filter with seed organisms should be employed.

3. Trickling Filter Plant Performance. The ammonia nitrogen conversion to nitrite and nitrate as a function of filter depth under different nitrogen loading conditions are shown in Figures 3, 4, and 5. It is apparent from these figures that as the filter depth increases, the NH_3-N concentration decreases with increasing the production of NO_2 and NO_3 in filter 1. The same relationship continuously existed in filter 2 until the net accumulation of nitrite became decreased. As a result, the rate of nitrate arose sharply but the removal of NH_3-N became insignificant.

The nitrogen loadings presently empolyed for the operation of both filters 1 and 2 are shown in tables 3 and 4. The nitrogen loading was calculated based on the kg NH_3-N used per tower surface area in m^2 per day (kg NH_3-N/m^2-day). According to tables 3 and 4, the nitrogen loading varied from 0.120 kg/m^2-day to 0.959 kg/m^2-day in filter 1 operation and varied from 0.016 kg/m^2-day to 0.530 kg/m^2-day in filter 2 operation. Since filter 2 treated the effluent of filter 1, the nitrogen loadings were lower.

The reduction of ammonia nitrogen in the plastic-media trickling filter system is certainly dependent upon the rate of nitrogen loading. Trickling filter 1 showed NH_3-N removal up to 86.4% for a nitrogen loading of 0.120 kg/m^2-day. As the loading was increased to 0.959 kg/m^2-day, the % NH_3-N was red-

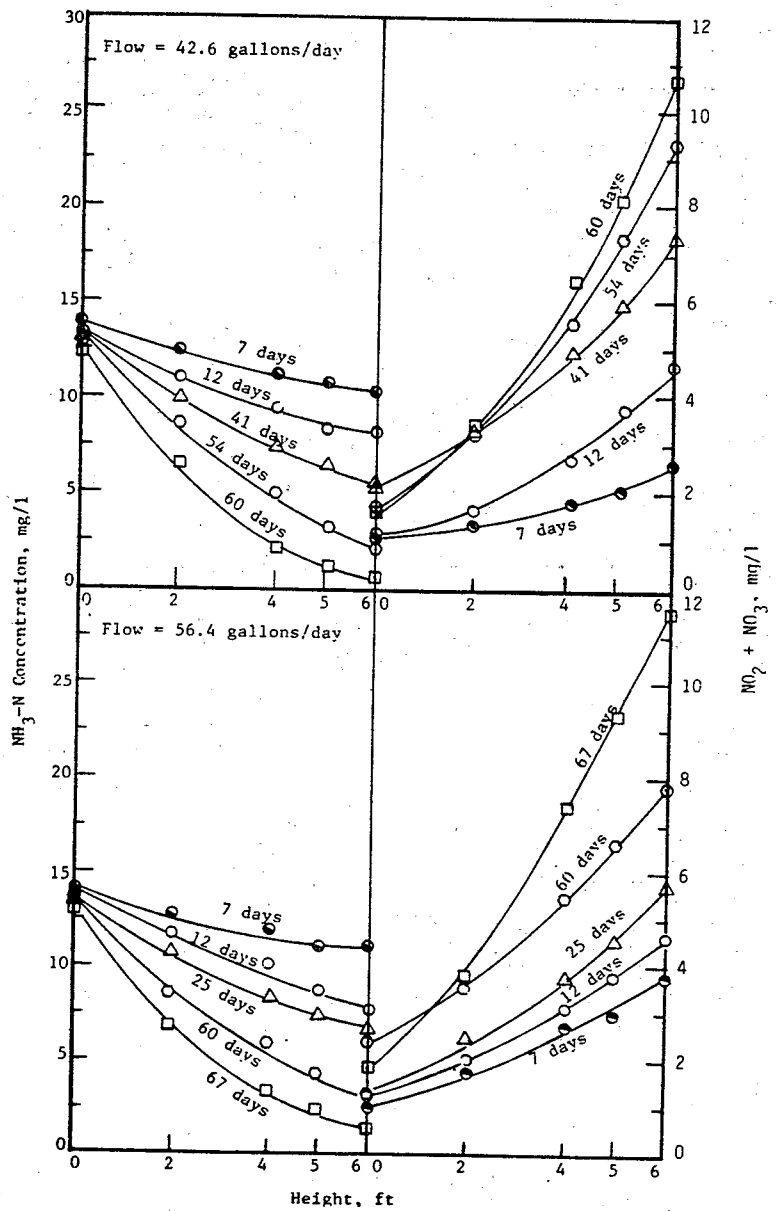


Figure 2. Trickling Filter Start-up

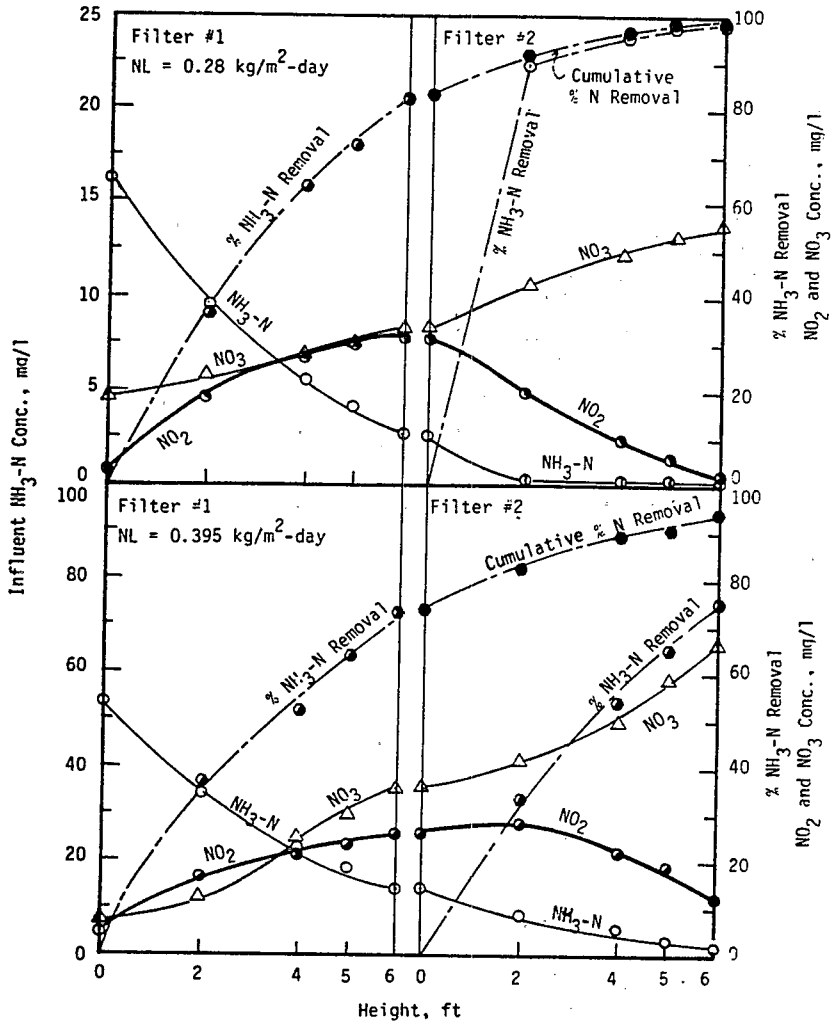


Figure 3. Ammonia Conversion as a Function of Nitrogen Loading at 0.28l and 0.395 kg/m²-day, respectively.

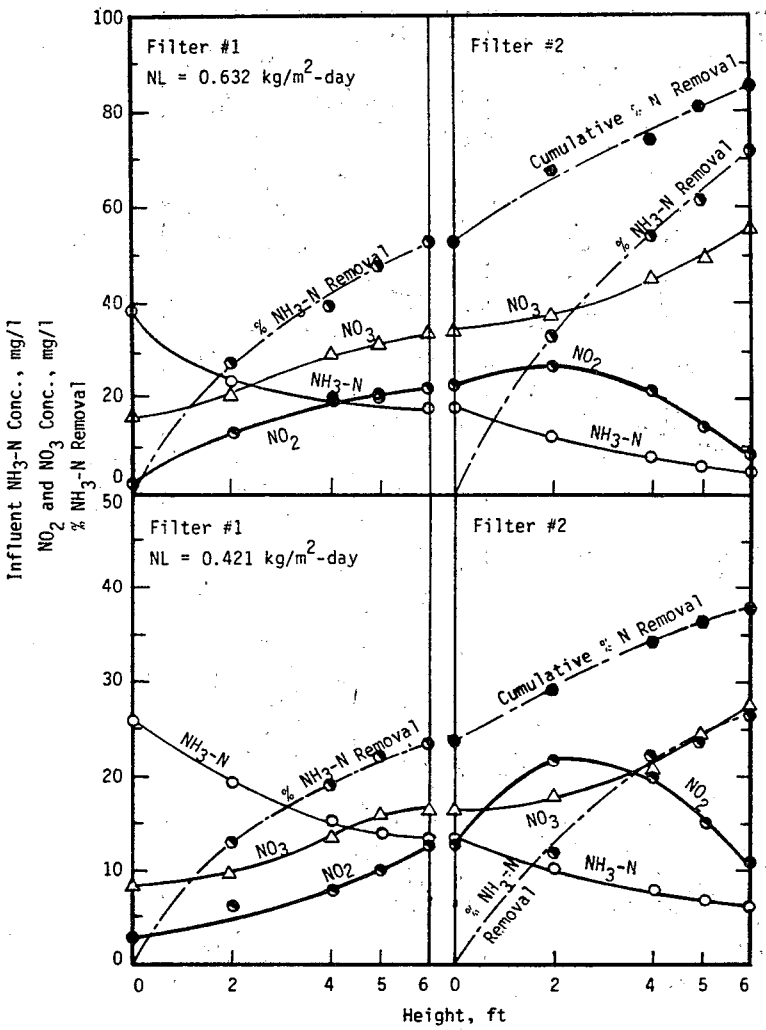


Figure 4. Nitrogen Conversion as a Function of Nitrogen Loading at 0.421 and 0.632 kg/m^2-day, respectively.

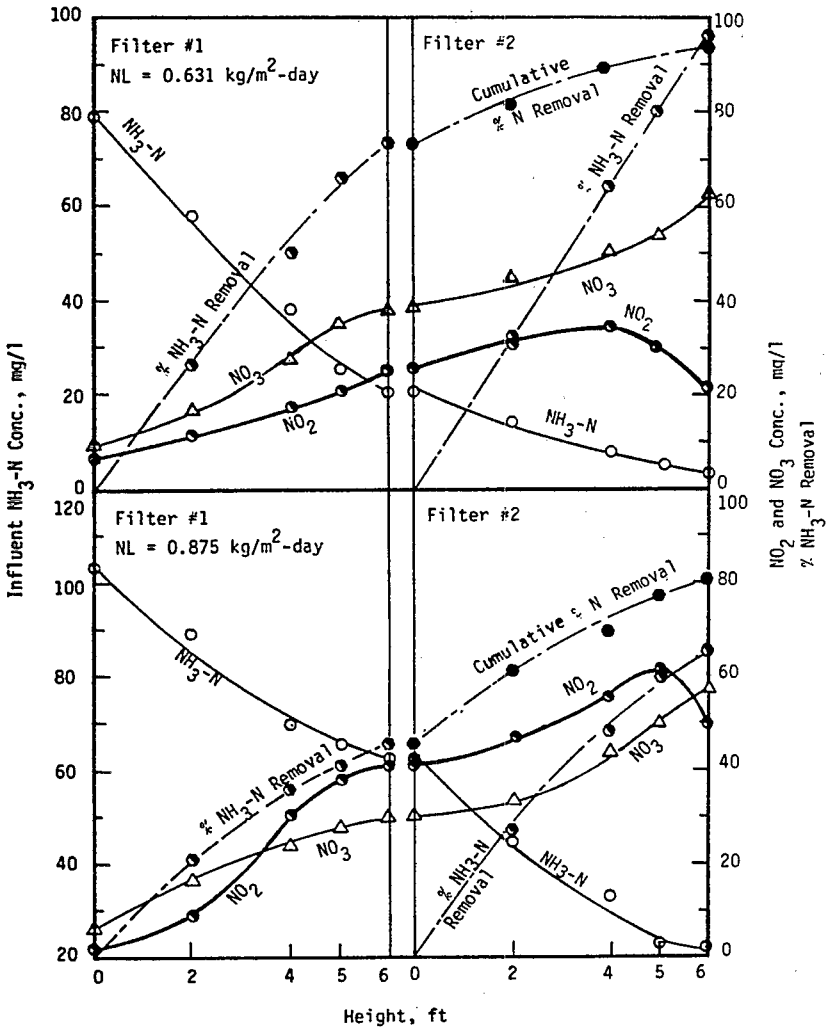


Figure 5. Ammonia Conversion as a Function of Nitrogen Loading at 0.631 and 0.875 kg/m^2-day, respectively.

Table 3. Trickling Filter 1 Effluent Characteristics

Nitrogen Loading Kg/m^2-day	Effluent NH_3-N mg/l	% NH_3-N Removal	NO_2 + NO_3 mg/l	pH	DO mg/l	TSS mg/l	VSS mg/l
0.959	41.4	44.2	41.75	8.0	4.1	7	4
0.875	62.0	43.3	58.18	7.8	3.1	10	9
0.837	24.7	50.6	32.93	7.5	6.2	9	7
0.630	29.3	63.0	63.12	7.8	4.6	6.6	4.9
0.632	15.0	62.0	25.43	7.7	7.2	11.3	8.6
0.578	15.0	67.0	31.02	7.8	4.0	3.6	2.3
0.421	7.0	73.1	20.65	7.8	7.3	5.3	3.0
0.361	5.8	79.3	32.23	7.8	4.7	4.6	2.7
0.395	14.0	74.1	55.18	7.7	4.3	13.7	7.0
0.349	7.7	82.2	27.55	8.0	4.2	3.0	2.5
0.311	3.3	82.9	16.23	7.7	7.2	20.0	11.0
0.280	2.5	83.1	16.37	8.2	4.6	2.7	2.2
0.129	1.8	86.7	15.6	7.7	4.5	3.3	2.6
0.120	1.7	86.4	13.02	8.1	4.6	6.2	2.5

Table 4. Trickling Filter 2 Effluent Characteristics

Nitrogen Loading kg/m^2-day	Effluent NH_3-N mg/l	% NH_3-N Removal	NO_2 + NO_3 mg/l	pH	DO mg/l	TSS	VSS mg/l
0.530	16.9	59.1	59.2	8.0	4.4	4.0	3.0
0.495	21.5	65.3	84.9	7.7	3.3	11.0	10.0
0.413	8.43	65.9	36.4	7.6	6.8	8.0	6.5
0.240	2.72	82.0	86.0	7.6	5.3	14.3	8.3
0.232	3.45	83.2	33.0	7.7	7.5	21.3	13.3
0.191	2.00	86.6	37.5	8.0	4.3	3.8	2.3
0.113	0.33	95.2	25.9	7.7	7.7	9.0	4.6
0.074	0.45	92.3	35.5	8.2	4.9	1.6	1.0
0.102	1.48	89.4	57.1	7.8	4.9	7.3	6.7
0.060	0.36	95.3	38.6	8.2	4.4	2.5	2.0
0.053	0.12	96.2	23.7	7.7	7.4	6.0	3.0
0.047	0.02	98.8	24.5	8.5	4.9	2.0	2.0
0.017	0.02	98.8	17.8	8.1	4.8	2.0	1.8
0.016	0.02	98.8	15.5	8.2	4.6	4.0	1.5

uced to 44.2. Trickling filter 2 achieved a higher nitrogen removal efficiency, ranging from 59.1% to 98.8%, due to a lower loading condition. According to the present study, two 6-foot filters connected in series are capable of obtaining an effluent NH_3-N Concentration less than 1.0 mg/l at a loading of 0.42 kg/m^2-day for first filter and of 0.113 kg/m^2-day for second filter. It is clear that the effluent NH_3-N remaining depends not only the nitrogen loading but also the initial concentration of NH_3-N in the wastewater. It is expected that the effluent NH_3-N concentration will be higher if the influent concentration of NH_3-N and the rate of nitrogen loading are higher.

Table 5 summarizes the nitrification data obtained from the present study and the other investigations using rotating biological contactors. The nitrogen loading used in this case was calculated based on pounds of NH_3-N applied per plastic media surface area in ft^2 per day (lb NH_3-N applied/1000 ft^2-day). Table 5 shows that the nitrification of secondary treatment effluent by trickling filter is comparable to or even better than that achieved by rotating biological contactors.

By comparing the pH and DO of the influent and effluent of filter 1, it is observed that both of these parameters fell slightly as the wastewater traveled through the filter. These results are normal because of nitrogen oxidation and destruction of bicarbonate alkalinity. The effect of nitrogen loading on alkalinity destruction will be discussed later in detail. Recovery of pH and DO was obatined after the wastewater passing through filter 2. The pH increase was very slight while the DO increase was more substantial. It seems obvious that the amount of oxygen dissolved in the wastewater is greater than the quanity required for biological nitrification in filer 2. No oxygen deficiency was found during the entire period of this study. Both effluents of filters 1 and 2 contained small concentrations of total and volatile suspended solids in accordance with tables 3 and 4.

The effect of filter height on the plant performance under various nitrogen loading conditions is shown in Figure 6. It can be seen in Figure 6 that for all loading conditions presently investigated the percentage of NH_3-N removal increases as the filter depth also increases. But such increase was reduced markedly when the nitrogen loading exceeded 0.8 kg/m^2-day in trickling filter 1. The results for trickling filter 2 are opposite. At the low nitrogen loading condition, the % removal of NH_3-N from the filter 1 effluent was only slightly affected by the filter depth measured at 2, 4, 5, and 6 feet below the top of plastic media. The influence of filter depth on trickling filter 2 performance becomes profoundly after the nitrogen load exceeding 0.5 kg/m^2-day.

Table 5. Nitrification of Secondary Effluent By Trickling Filter and Rotating Biological Contactors

Nitrogen Loading Rate lb NH_3-N Applied / 1000 ft^2- day			Ammonia Nitrogen Reduction (%)		
Trickling Filter (a)	Rotating Biological Contactor (b)[9]	(c)[10]	Triclking Filter (a)	Rotating Biological Contactor (b)	(c)
0.067			99.9		
0.072		0.08	99.8		94.4
0.176		0.13	99.3		70.3
0.185		0.18	98.9		83.1
0.198	0.20	0.21	98.9	95.0	88.3
0.206			97.2		
0.225		0.22	98.4		91.7
0.241			98.7		
0.332			95.6		
0.362			83.2		
0.364	0.40	0.42	93.1	90.0	76.8
0.482	0.50	0.48	80.3	80.0	83.1
0.504			77.1		
	0.60			90.0	
	0.61			65.0	
	0.76			13.0	
	0.78			73.0	
		0.83			70.3
	1.10			25.0	
	1.29			35.0	

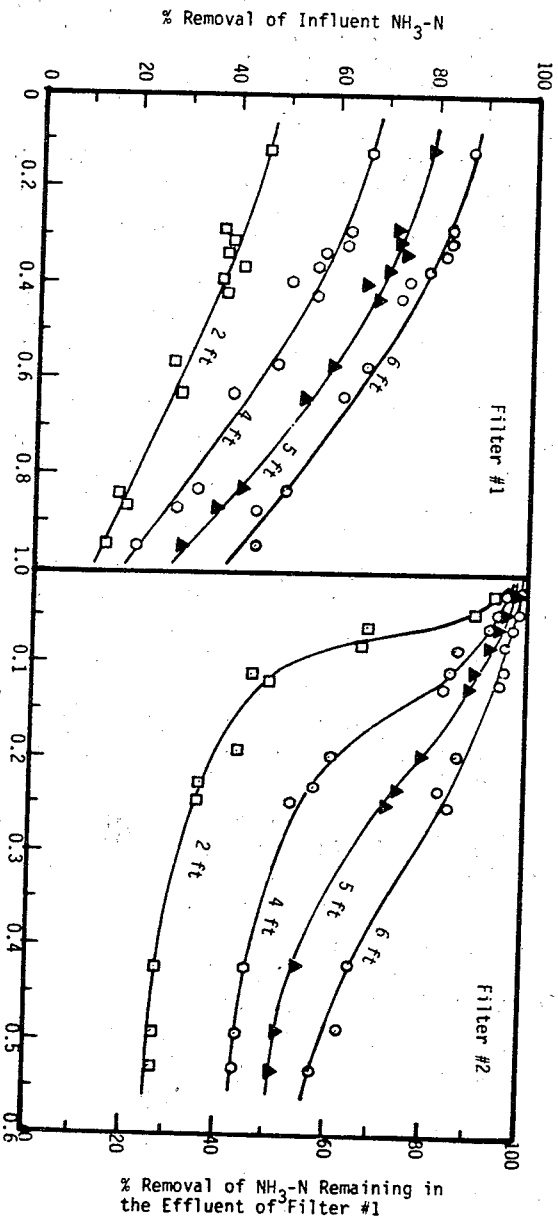


Figure 6. Relationship Between Nitrogen Loading and % NH_3-N Removal At Different Filter Depths

4. Alkalinity Requirement. The pound of alkalinity destoryed per pound of ammonia nitrogen oxidized to nitrate normally equals to 7.2 (11). A plot of pound of alkalinity consumed per pound of NH_3-N removed is shown in Figure 7. It is clearly shown in Figure 7 that the alkalinity requirement for trickling filters 1 and 2 is different. And the normal ratio of 7.2 was not observed.

Trickling filter 1 shows a ratio of approximately 6.7 up to a nitrogen loading of 0.50 kg/m^2-day. As the loading exceeded the above mentioned value, the pound of alkalinity utilized per pound of ammonia nitrogen oxidized became smaller. For instance, at the nitrogen loading of 0.959 kg/m^2-day, the ratio is 4.6 Trickling filter 2 shows the same resulting curve as filter 1. According to Figure 7, the ratio varied from 7.8 to 4.25 as the nitrogen loading increased from 0.016 kg/m^2-day to 0.53 kg/m^2-day. The ratio decrease became so apparent when the nitrogen loading exceeding 0.25 kg/m^2-day.

It also appeared that the effect of nitrogen loading on alkalinity demand for nitrification in the trickling filter system was significant, in particular, when the nitrogen load exceeding 0.5 and 0.25 kg/m^2-day in the first and second filter, respectively. Although the organic content in the wastewater was low (<25 mg/l as BOD), ammonia nitrogen was utilized by both nitrifying bacteria and heterotrophic organisms due to the fast feed rate. This is the reason to explain why the alkalinity demand per unit amount of nitrogen consumed was low at high nitrogen loading condition.

5. Sludge Settling Characteristics. Nitrifying organisms produced from the fixed-film trickling filter system have an excellent property in settling. Although the SVI was high between 199 and 319 during the strat-up period, its value varied only slightly around 110 at all times after the system reached the steady-state condition. Normal sludge floc is shown in Figure 8. The size of the floc was large and it can be separated from liquid phase quickly. However, regular microscopic examination of filter sludge occasionally found that long-length filamentous microorganisms existed. It is believed that these filaments are transferred from the effluent of the activated sludge plant instead of being developed from the trickling filters. The morphological structure of the long-length filaments is shown in Figure 9.

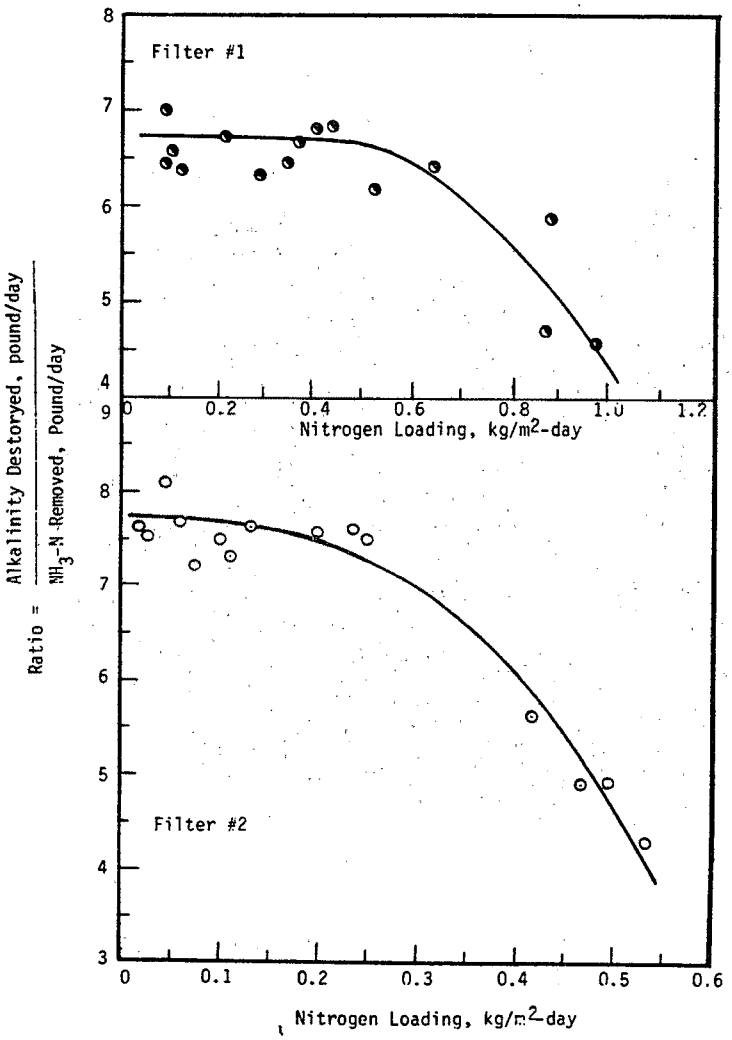


Figure 7. Alkalinity Destruction as a Function of Nitrogen Loading

Figure 8. Normal Trickling Filter Sludge (200 X)

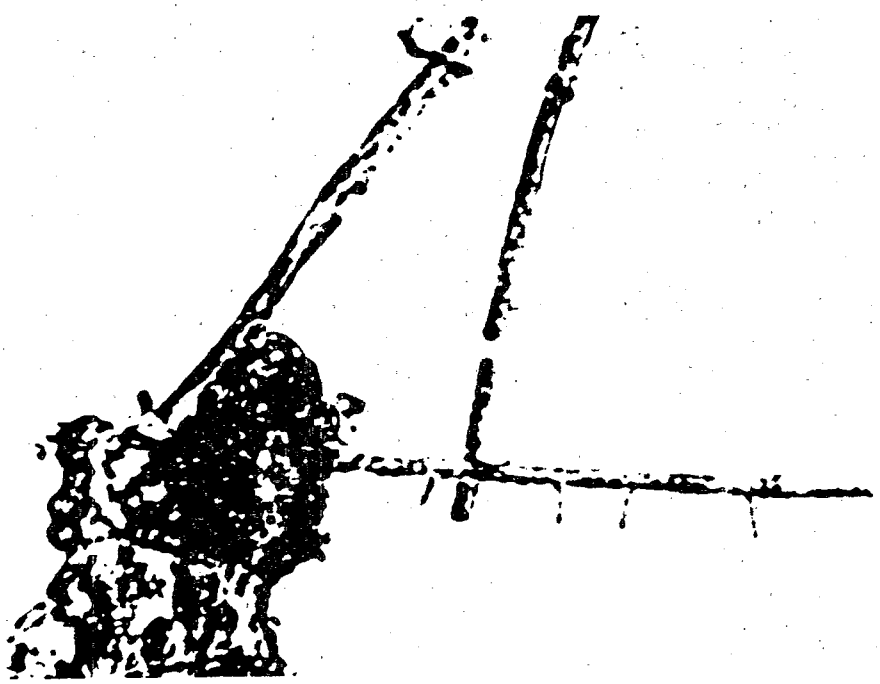


Figure 9. Filamentous Growth in Trickling Filter Sludge (200 X)

CONCLUSIONS

Biological nitrification of secondary effluent by a two-stage trickling filter was throughly investigated under the optimum pH and temperature conditions. It was found that the efficiency of the trickling filter plant was a function of influent nitrogen concentration, nitrogen loading, and filter depth. The quantity of ammonia nitrogen removed was higher in filter 1 than filter 2. The continuously oxidized ammonia nitrogen resulted in the accumulation of nitrite and nitrate in the system, however, the production of nitrite started to decrease with increasing the nitrate when ammonia nitrogen remaining in the wastewater approached to its lowest level. More than 93% of ammonia nitrogen can be removed by passing it through two six-foot plastic media trickl-filters, operated at the nitrogen loading equal to 0.63 kg/m^2-day in the first filter and 0.24 kg/m^2-day in the second filter.

Both pH and DO decreased and increased after the wastewater traveled through filter 1 and filter 2, respectively. Dissolved oxygen never became the growth-limiting factor because its concentration was over 3.0 mg/l. The effluent total and suspended solids were extremely low and the sludge settleability was very high with an averaged SVI = 110. Filamentous microorgansims was seldom found in the sludge and they were probably transferred to the filter from the effluent of the activated sludge plant. That is why the outgrowth of filamentous microorganisms never occurred.

The alkalinity requirement for fixed-film biological nitrification is somewhat different from that observed from suspended growth systems. The quantity of alkalinity destoryed per unit of ammonia nitrogen removed is always below normal value of 7.2 in both filters presently investigated if the nitrogen loading was kept to exceed 0.25 kg/m^2-day. Both nitrifiers and heterotrophic microrganisms play an important role in nitrogen assimilation when the feed rate and the nitrogen load are high. This explains why the alkalinity demand for biological nitrification is lower under the above mentioned condition.

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PART VIII: INDUSTRIAL WASTEWATER TREATMENT

UPGRADING SLAUGHTERHOUSE EFFLUENT WITH ROTATING BIOLOGICAL CONTACTORS

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INTRODUCTION

The purpose of this work was to extend the potentially attractive rotating biological contactor (RBC) process to upgrade treated slaughterhouse wastewater effluent from a biological tower. The research was primarily concerned with demonstrating the RBC potential for removing biochemical oxygen demand (BOD) from low-temperature wastewater. Attention was concentrated on the time to reach process stability and treatment performance with low-temperature wastewater.

The RBC-pilot plant, described more completely below, receives a constant wastewater flow of 1.6 l/min. of settled effluent from a tower trickling filter (bio-tower). The organic loading to the pilot plant varies over time; from a soluble BOD_7 of 50 mg/l to as high as 700 mg/l. The higher organic loading does not negatively affect RBC-treatment performance. In fact, percent removal of BOD increases with increasing organic loading.

Process stability measured as percent BOD removed in $10^{o}C$ wastewater, was reached in approximately three weeks after start-up. The bio-tower preceeding the RBC is

considered a "roughing" process as it reduces the organic loading on the RBC to a level that allows "optimal" removal. This combination, bio-tower and RBC, therefore, enables the treatment plant to meet new secondary treatment requirements and other possibly more stringent requirements adoptable in the future.

According to the literature, the number of treatment flowsheets that can be derived by combining various biological treatment processes is nearly endless. Combinations of trickling-filter and activated-sludge processes are the more common flowsheets. These processes have be used successfully for a number of years for the treatment of all types of wastewater, especially combined domestic and industrial wastewater (1).

The process microbiology for combined biological treatment processes is essentially the same as for the individual processes. The biological activity in the bio-tower, the "roughing" filter in this research, will be somewhat different from the RBC because of the higher shearing action resulting from the hydraulic flowrates applied to the tower. The bio-tower moreover acts to reduce the organic loading on the RBC, making nitrification a possibility, expecially during periods with higher wastewater temperatures, i.e.; above 10°C.

The bio-tower effluent is piped to an adjoining fjord with the outfall-pipe extending to a depth of 43 m. The fjord is showing signs of deterioration with algae growth and anoxic zones due to the total communal load of domestic and industrial wastewater. It is therefore encouraging that the RBC-pilot plant performs well as this process may be the future mode of "polishing" the existing bio-tower effluent.

INSTALLATION AND START-UP

Figure 1 shows the location of the plant. The present effluent is piped through two 0.16 m diameter pipes, a length of 240 m along the side-slope of the fjord, to a depth of 43 m. A schematic of the treatment plant is shown in Figure 2.

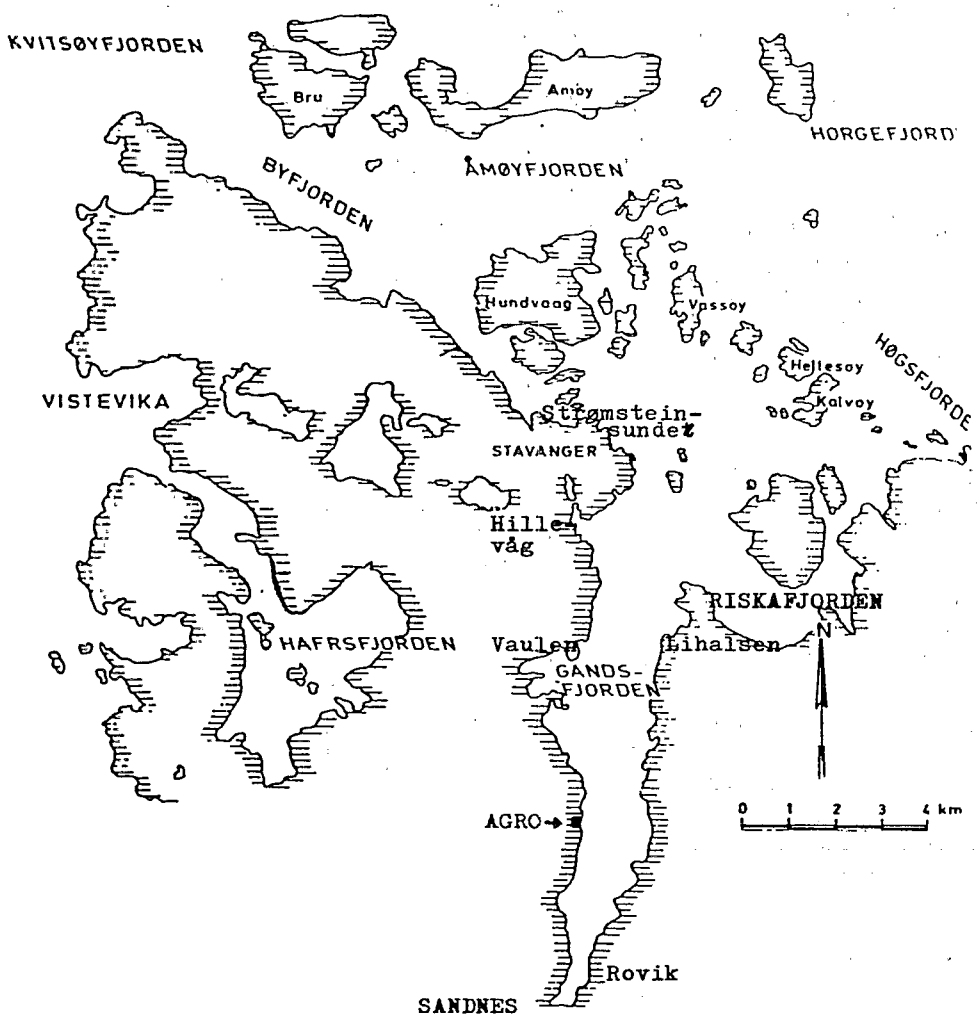

Figure 1. The fjord-system surrounding the cities of Sandnes and Stavanger in the south-west of Norway. The slaughterhouse is situated at AGRO, in the lower portion of the figure.

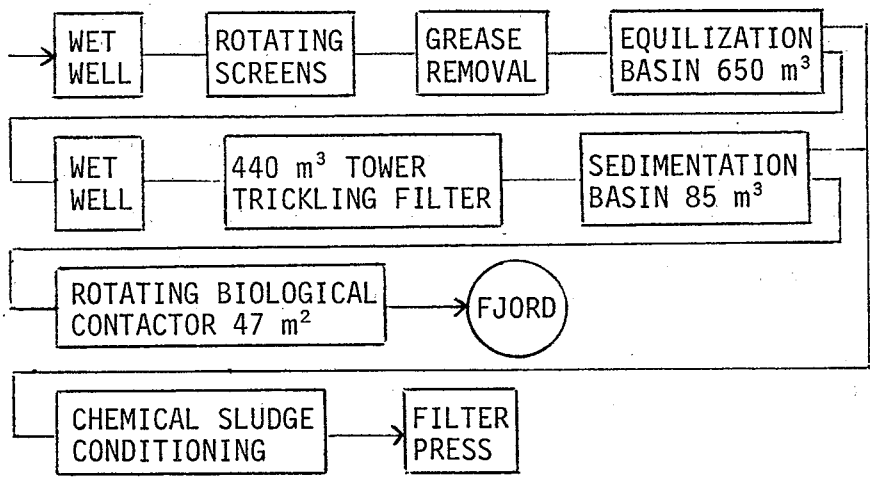

Figure 2. Slaughterhouse wastewater treatment flow-diagram

The slaughterhouse operates on one eight-hour work shift per day, beginning at 7:00 a.m. Treated effluent from the 440 m^3, 7.8 m high plastic packed bio-tower is recycled at a rate of 50% during slaughterhouse production hours, and 100% when there is no flow from the equilization basin. The recycling of treated effluent is performed in order to dilute the raw wastewater from the slaughterhouse before it reaches the bio-mass in the tower, and to ascertain fluid to the bio-mass at any one time, especially during off-production hours at the slaughterhouse, to prevent dry-out of the microorganisms.

The pilot plant module is a 0.92 m diameter four-stage RBC unit mounted in a fiberglas tank; S5 Rotordisk, manufactured by CMS Equipment Limited, Mississauga, Ontario, Canada. Each stage is 0.3 m long and the total contact surface area for the entire unit is 47 m^2 (500 ft^2). The four-stage rotorzone volume is 0.41 m^3 (14.5 ft^3) and gives a theoretical detention time of 4.3 hours at a manufacturer recommended wastewater flow of 2.3 m^3/d, equivalent to a hydraulic loading of 0.05 $m^3/m^2 \cdot d$.

Approximately 40% of the RBC surface area is submerged in wastewater and the rotational speed is three revolutions each minute giving a peripheral speed of 0.14 m/s. The S5 Rotordisk is a package wastewater treatment plant and

contains in addition to the four rotorzones, a 1.4 m^3 primary clarifyer and a 0.6 m^3 final clarifyer. The whole unit is enclosed. Figure 3 gives the plan view and a sectional side view of the RBC-pilot plant.

After start-up, time was allowed for a biological slime layer to develop on the rotating media. The average wastewater temperature during the start-up period was approximately 10°C throughout the RBC. The hydraulic loading was constant at 0.05 $m^3/m^2 \cdot d$, whereas the organic loading varied substantially over time with an average total BOD_7 of 645 mg/l, equivalent to an organic loading of 32 g total $BOD_7/m^2 \cdot d$.

Table I and Figure 4 indicate that process stability was reached in approximately three weeks. In fact total BOD removal was 71% and total chemical oxygen demand removal (COD) was 61% after 16 days. Obviously, this period was too short for nitrification to develop and hence, removal of ammonia nitrogen (NH_3) was not evident. The 10°C wastewater temperature and high organic loading did not exactly provide optimal conditions for the nitrifying organisms. The pH was approximately 7.5 and the alkalinity was 50 mg/l as $CaCO_3$ or higher, over the same period.

SAMPLING AND ANALYSIS

Several series of experiments have been conducted during the weeks after start-up. The wastewater temperature through the RBC has been relatively cold; approximately 10°C. Grab samples of influent to the bio-tower, of influent to the RBC, of wastewater from the four RBC-stages, and from the RBC-effluent were collected at various times in order to determine changes in wastewater characteristics through the treatment system. Grab samples were periodically coupled with flow-proportioned composite samples collected either manually or by an ISCO Modell 2100 automatic wastewater sampler. Wastewater flow, temperature, pH and dissolved oxygen (DO) were measured at the plant during sampling. The choice of laboratory parameters varied and included alkalinity, nitrogen, phosphorous, BOD, COD and total organic carbon (TOC). Unless otherwise stated, analyses were conducted on filtrate passing a 1 µm poresize glasfibre filter.

The TOC analyses were conducted on a Beckman, Modell 915-B Total Organic Carbon Analyzer. Continuous BOD versus

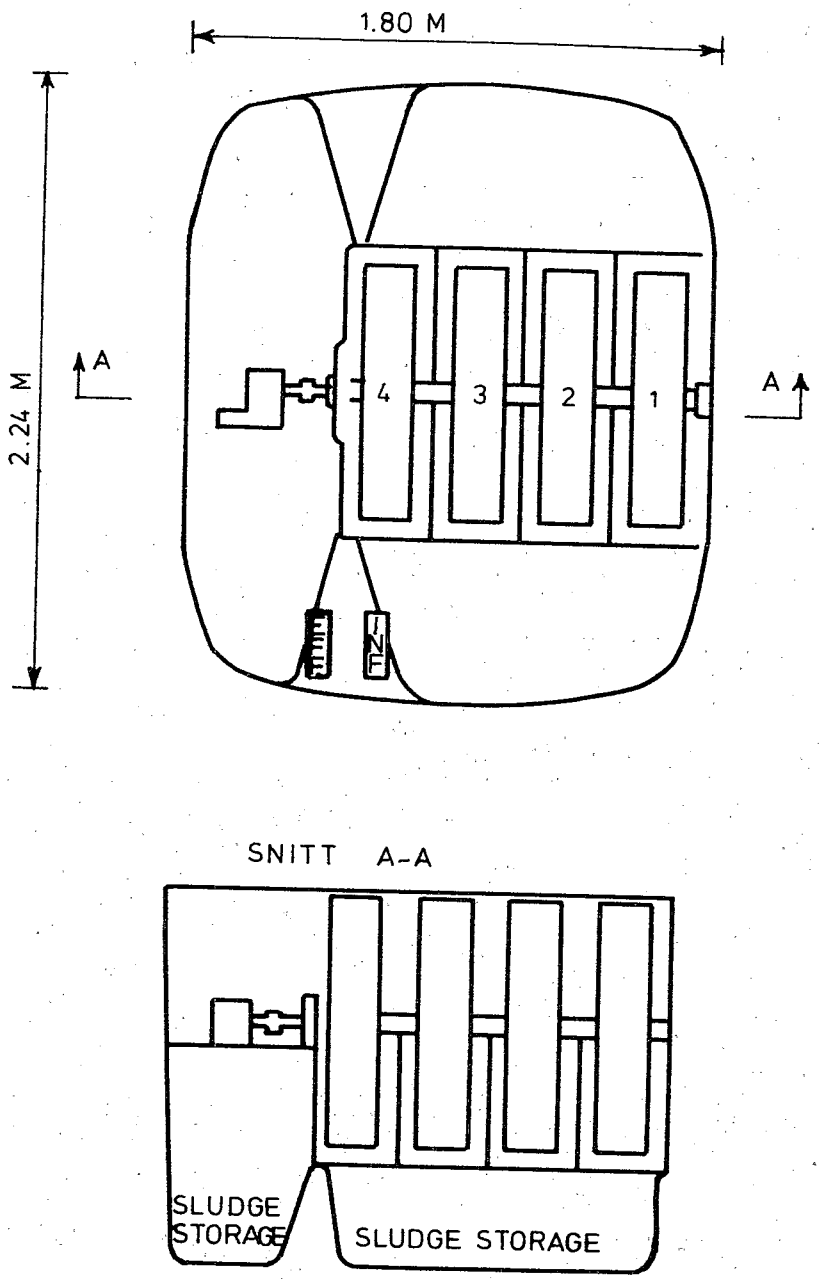

Figure 3. Plan and sectional side view of the RBC, S5 Rotordisk.

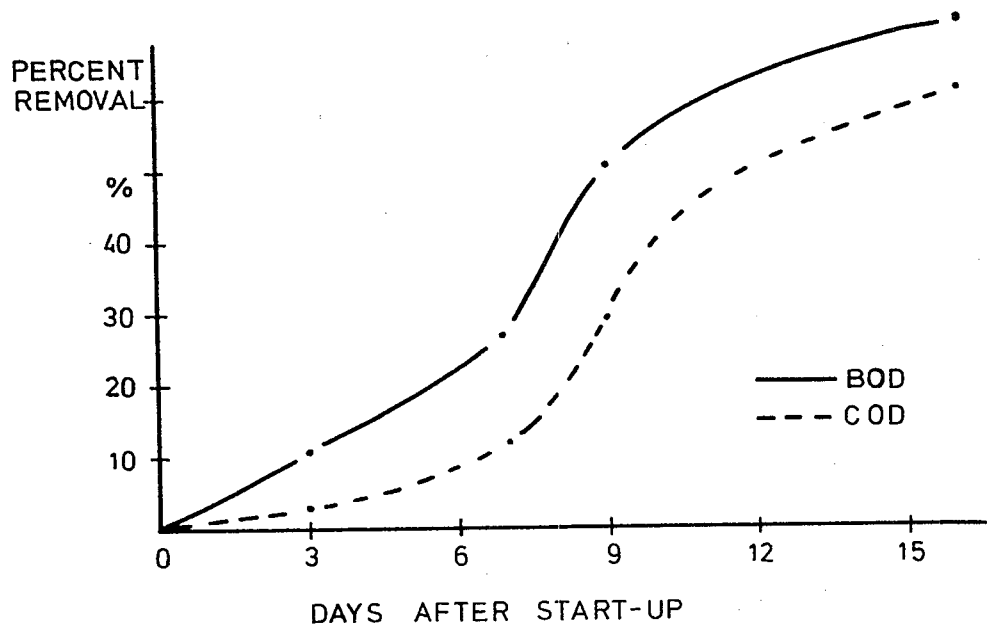

Figure 4. Total BOD_7 and COD removal during start-up of the RBC.

Table I

RBC Treatment Characterictics with $10^{o}C$ Wastewater.

Days after start-up	$TBOD_7$ (mg/l)			TCOD (mg/l)		
	Inf.	Eff.	%R	Inf.	Eff.	%R
3	295	263	11	568	550	3
7	525	385	27	785	690	12
9	780	382	51	900	640	29
16	630	185	71	971	384	61

time curves were developed on an automatic instrument by use of electrolytic respirometry; Voith Sapromat, Model C 12, J.M. Voith GmbH, 7920 Heidenheim, West Germany. A schematic diagram of a measuring unit is shown in Figure 5. Each measuring unit comprises one reaction vessel, one oxygen generator and one pressure indicator which are interconnected by plastic hoses. The sealed measuring system is not affected by barometric air pressure fluctuations.

The required oxygen for the microorganisms is at any time available in the electrolytic cell and is always supplied in sufficient quantity to the sample of wastewater to be analyzed. The BOD value which can be measured is limited as the maximum oxygen demand of the sample may not exceed 90 mg/l·h. If BOD_5 values higher than 3000 mg/l are encountered, it is good practice to dilute the sample. Therefore, in contrast to the conventional BOD dilution method, a genuine respiratory process takes place in the Sapromat.

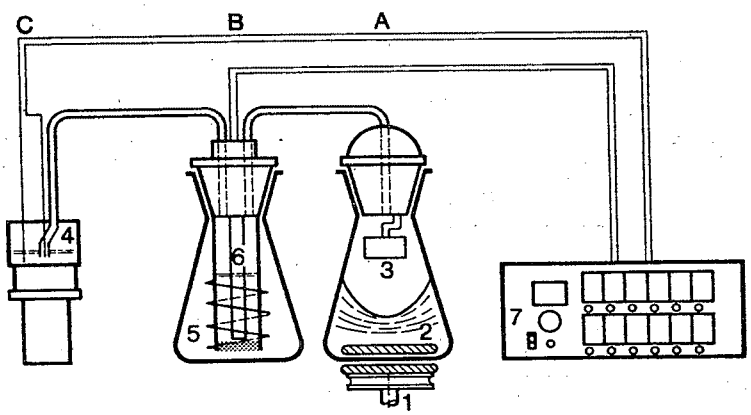

Figure 5. Schematic diagram of a Sapromat measuring unit where A = Reaction vessel, B = Oxygen generator, C = Pressure indicator, 1 = Magnetic stirrer, 2 = Sample, 3 = CO_2 absorber, 4 = Pressure indicator, 5 = Electrolyte, 6 = Electrodes, 7 = Measuring and control unit with digital printer

RESULTS AND DISCUSSION

A typical treatment performance of the bio-tower and RBC system, four and five weeks after start-up, is summarized in Tables II and III respectively. The RBC pilot plant reaches equilibrium with respect to organic removal approximately three weeks after start-up. However, as would be expected for the 10°C wastewater, nitrification was not apparent until the fifth week of operation. Impending nitrification is illustrated by a small production of nitrites (NO2) and nitrates (NO3) as in Figure 6. The RBC pilot plant has not experienced problems with low concentrations of dissolved oxygen (DO). Figure 7 is typical in this respect, indicating adequate oxygen mass transfer in all four RBC-stages.

Four Weeks After Start-up

The soluble (filtered) long-term BOD concentrations obtained from the bio-tower and RBC pilot plant during a typical slaughterhouse production day are illustrated in Figure 8. The analyses were performed on composite afternoon samples, each comprised of five half-hourly grab samples. These long-term BOD curves are especially illustrative in the case of the RBC-effluent where the effect of nitrification showed up very clearly for the non-inhibited nitrogenous BOD. The Sapromat oxygen uptake reaction was performed on non-diluted wastewater which, no doubt, was of great importance for the slow-growing nitrifiers. The standard dilution BOD-method often retards nitrification because of a population decrease of the nitrifying bacteria (2).

The rate of nitrification, and therefore the rate of growth and length of generation times of the nitrifying organisms, is affected by several environmental factors including temperature, pH and dissolved oxygen. Borchardt (3) estimated the rate of nitrification at 9°C to be about 50% of the rate at 20°C. The RBC did not provide a long enough sludge-age after four weeks of operation to provide nitrification at 9°C. The 20°C water-bath in the Sapromat coupled with six days of incubation provided such a sludge-age for the RBC-effluent sample (Figure 8). The pH of the wastewater was normally above 7.0.

Successful removal of carbonaceous material expressed as BOD, COD or TOC is apparent from Figure 9 and Table II. The hydraulic loading was constant at 0.05 $m^3/m^2 \cdot d$ which yielded

Table II

Treatment Plant Performance Four Weeks After Start-up. Wastewater temperature through the RBC ≅ 9°C.

Sample Time (o'clock)	Sample Station	Alkalinity $mgCaCO_3/l$	BOD_7	COD	TOC	TKN	NO2+NO3
			mgO_2/l		mgC/l	mgN/l	
07:30	Tower-IN	25	-	112	54	9.8	0.1
"	RBC-IN	135	45	86	32	29.2	1.0
"	RBC-1	290	12	125	60	30.4	0.2
"	RBC-EFF	265	10	128	72	61.1	0.2
14:30	Tower-IN	85	700	1032	290	84.7	0.1
"	RBC-IN	220	390	560	115	70.8	0.2
"	RBC-1	225	62	109	47	48.3	0.2
"	RBC-EFF	205	34	74	41	47.4	0.4
20:30	Tower-IN	100	-	566	154	43.0	0.0
"	RBC-IN	240	-	360	107	70.3	0.1
"	RBC-1	285	51	152	87	41.7	0.1
"	RBC-EFF	245	40	117	67	56.6	0.2

Table III

Treatment Plant Performance Five Weeks After Start-up.

Sample Time (o'clock)	Sample Station	T°C	DO mg/l	pH	Alka-linity $mgCaCO_3/l$	COD $\frac{mgO_2}{l}$	TP	PO_4	TKN	NO2+NO3
							mgP/l		mgN/l	
07:30	Tower-IN	-	-	-	65	640	7.2	4.7	62.1	0.1
"	RBC-IN	13	0.5	-	80	128	6.3	5.7	66.9	0.0
"	RBC-1	11	6.2	6.9	60	100	5.6	5.1	42.0	3.3
"	RBC-2	11	7.0	7.0	65	103	5.5	4.8	45.5	2.9
"	RBC-3	10.8	7.0	7.1	50	117	5.6	4.8	45.8	4.8
"	RBC-4	10.5	7.5	7.1	45	94	5.5	4.9	38.5	5.6
"	RBC-EFF	-	-	-	45	81	5.4	4.8	38.3	5.7
09:30	Tower-IN	-	-	-	140	1565	16.1	10.5	131.8	0.2
"	RBC-IN	-	-	-	70	139	6.7	5.7	61.9	4.1
"	RBC-1	-	-	-	65	100	5.9	5.3	44.2	3.3
"	RBC-2	-	-	-	75	89	5.9	5.3	44.2	2.3
"	RBC-3	-	-	-	55	75	5.8	5.1	42.3	4.8
"	RBC-4	-	-	-	47	82	5.6	5.1	39.7	5.6
"	RBC-EFF	-	-	-	50	93	5.6	5.0	38.3	5.7

Table III continued on next page

Table III (continued)

Sample Time (o'clock)	Sample Station	T°C	DO mg/l	pH	Alka-linity $mgCaCO_3/l$	COD $\frac{mgO_2}{l}$	TP	PO_4	TKN	NO2+NO3
							mgP/l		mgN/l	
11:30	Tower-IN	-	-	7.1	130	1725	17.1	11.7	144.7	0.3
"	RBC-IN	14	1.8	-	160	284	7.9	6.2	75.1	2.7
"	RBC-1	11	5.2	6.9	70	100	6.1	5.5	52.8	3.8
"	RBC-2	11	6.2	7.1	65	89	5.8	5.2	46.5	3.3
"	RBC-3	10.8	7.5	7.1	55	81	5.7	5.2	43.1	5.2
"	RBC-4	10.5	7.8	7.1	49	105	5.6	5.2	43.9	5.9
"	RBC-EFF	-	-	-	55	99	5.6	5.1	40.6	5.9
13:30	Tower-IN	-	-	-	175	1725	15.1	10.0	129.7	0.3
"	RBC-IN	-	-	-	255	520	9.5	7.3	95.3	1.1
"	RBC-1	-	-	-	90	145	6.3	5.2	53.0	3.6
"	RBC-2	-	-	-	80	117	6.1	5.1	45.7	4.1
"	RBC-3	-	-	-	65	111	6.1	5.1	45.2	5.3
"	RBC-4	-	-	-	55	114	5.9	4.8	44.8	6.2
"	RBC-EFF	-	-	-	55	109	5.9	4.8	42.9	6.1

Table III continued on next page

Table III (continued)

Sample Time (o'clock)	Sample Station	T°C	DO mg/l	pH	Alka-linity $mgCaCO_3/l$	COD $\frac{mgO_2}{l}$	TP	PO_4	TKN	NO2+NO3
							mgP/l		mgN/l	
15:30	Tower-IN	-	-	-	150	1490	18.1	12.0	163.7	0.3
"	RBC-IN	14.5	2.2	7.7	295	454	9.6	7.0	108.6	0.4
"	RBC-1	10.5	3.3	7.3	150	312	7.2	5.9	62.7	1.6
"	RBC-2	10.0	5.4	7.3	105	239	6.4	5.3	56.6	3.5
"	RBC-3	9.5	6.5	7.4	85	178	6.4	5.2	55.3	4.8
"	RBC-4	9.5	7.1	7.3	75	117	6.2	5.2	46.1	6.3
"	RBC-EFF	-	-	-	60	131	5.9	5.1	44.4	6.3
17:30	Tower-IN	-	-	-	115	883	13.3	8.8	105.8	0.2
"	RBC-IN	-	-	-	330	623	10.6	8.0	129.7	0.3
"	RBC-1	-	-	-	170	195	7.2	5.9	175.4	0.8
"	RBC-2	-	-	-	140	153	6.7	5.5	65.3	2.3
"	RBC-3	-	-	-	115	105	6.3	5.3	64.1	4.3
"	RBC-4	-	-	-	100	113	6.1	5.2	63.1	5.2
"	RBC-EFF	-	-	-	75	88	6.0	5.1	51.9	5.9

Table III continued on next page

Table III (continued)

Sample Time (o'clock)	Sample Station	T°C	DO mg/l	pH	Alka-linity $mgCaCO_3/l$	COD $\frac{mgO_2}{l}$	TP	PO_4	TKN	NO2+NO3
							mgP/l		mgN/l	
19:30	Tower-IN	-	-	-	90	278	10.2	6.6	53.2	0.1
"	RBC-IN	14.0	0.5	7.7	360	600	9.1	6.1	118.7	0.3
"	RBC-1	11.5	3.7	7.5	195	225	7.6	6.1	71.8	0.3
"	RBC-2	11.0	5.0	7.6	190	203	7.2	6.1	69.0	1.2
"	RBC-3	11.0	6.0	7.7	170	173	6.9	5.6	56.9	2.8
"	RBC-4	10.8	6.5	7.7	150	150	6.9	5.4	54.7	3.8
"	RBC-EFF	-	-	-	110	117	6.5	5.5	52.8	5.2

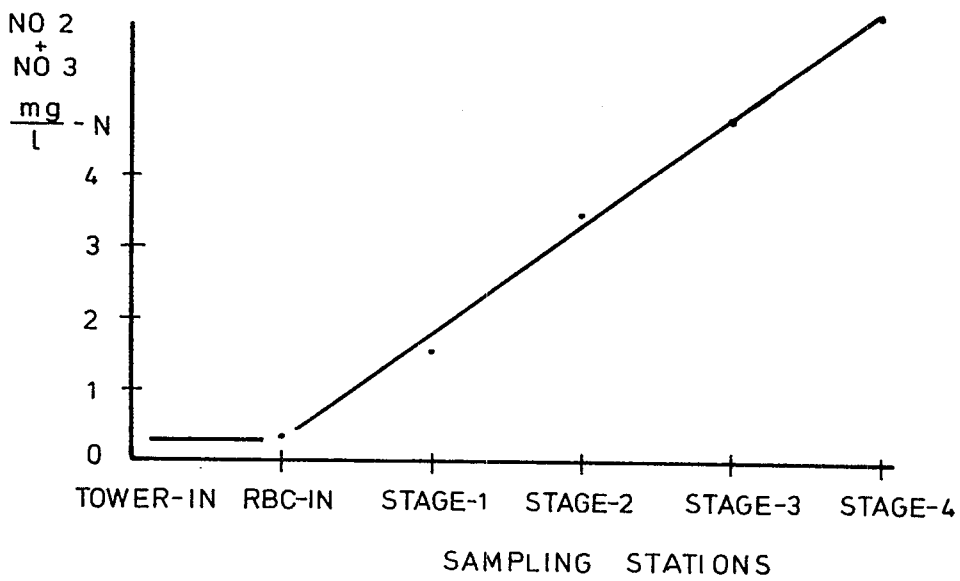

Figure 6. Impending nitrification depicted by a small production of nitrite and nitrate five weeks after start-up.

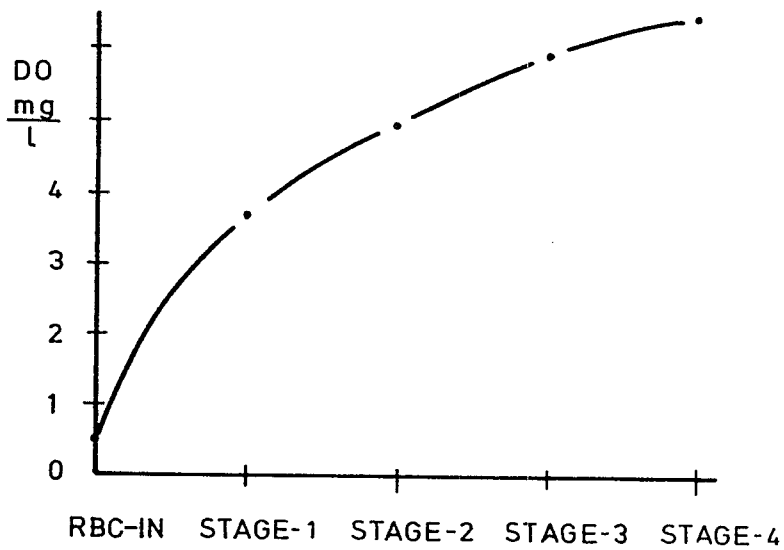

Figure 7. Dissolved oxygen through the RBC system.

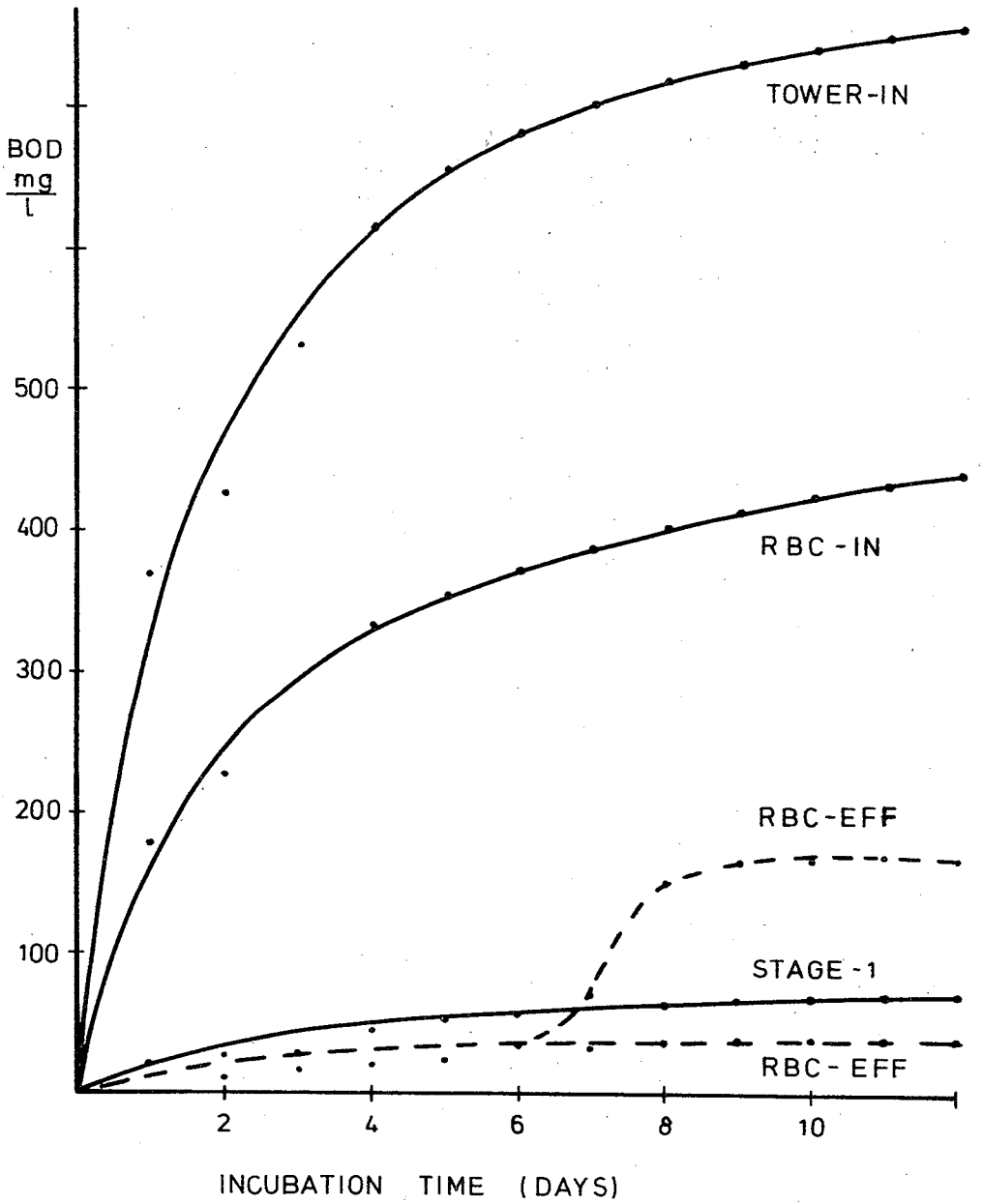

Figure 8. Long-term soluble BOD analyzed on the Sapromat. Nitrification is evident in the one RBC-effluent sample where nitrification is not inhibited.

an organic loading of 17.8 g $BOD_5/m^2 \cdot d$ and 19.5 g $BOD_7/m^2 \cdot d$. Maximum recommended loading values in Norway are 15 g $BOD_7/m^2 \cdot d$ at $10^{\circ}C$ for 85% BOD_7-removal, and a maximum of 7.5 g $BOC_7/m^2 \cdot d$ if nitrification is intended (4).

The removal of soluble BOD was more than 90% in the RBC pilot plant. The effluent BOD_5 and BOD_7 were approximately 25 mg/l and 35 mg/l respectively, four weeks after start-up. To give an indication of the solid separation efficiency in the RBC-final clarifier, the total (unfiltered) BOD_5 and BOD_7 were 38 mg/l and 46 mg/l respectively for the situation

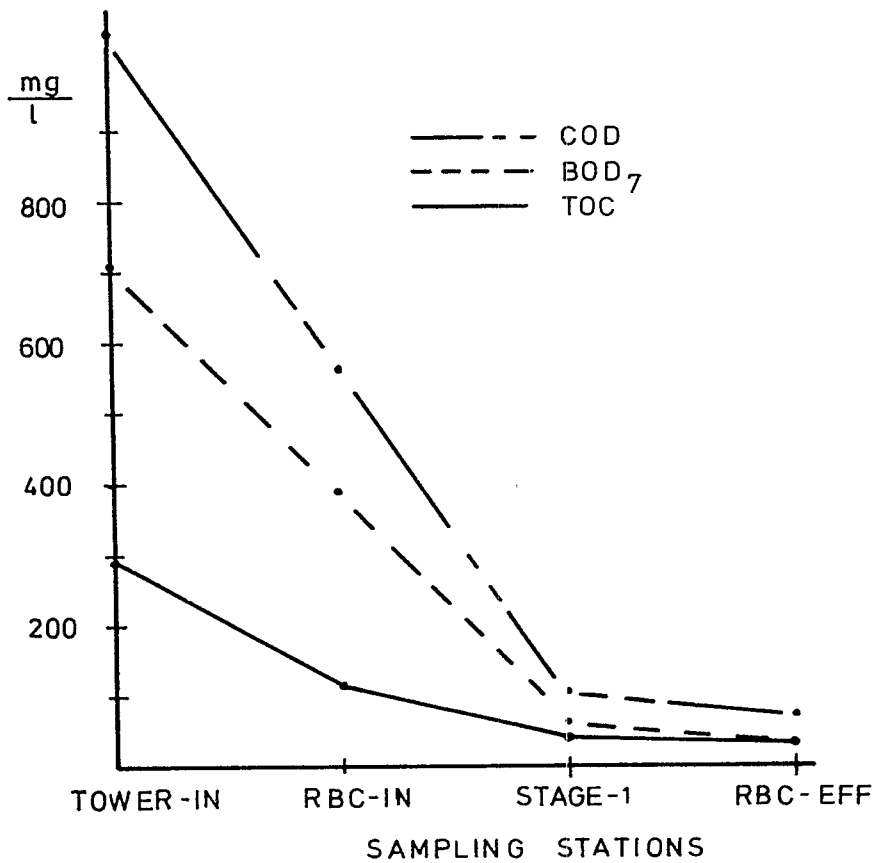

Figure 9. Removal of substrate in the tower trickling filter and rotating biological contactor four weeks after start-up.

depicted in Figures 8 and 9. The evidence furthermore indicated that the soluble BOD removal percentage in the RBC pilot plant increased with increasing BOD concentration in the influent.

Five Weeks After Start-up

The data presented in Table III and Figure 6 suggest impending nitrification. The increasing concentration of NO2+ NO3 through the RBC pilot plant was consistent during the 24-hour period making up the data. This was not unwarranted as Figure 8 revealed the presence of nitrifiers in the RBC-effluent four weeks after start-up. The BOD-test was performed on fiberglass-filtered, undiluted and nonseeded wastewater and, as mentioned earlier, the impending nitrification would probably not have been detected by a standard BOD-dilution test. In addition to dilution of the sample, the standard BOD test also incorporates seeding with raw wastewater containing heterotrophic bacteria thereby possibly negating nitrification.

The principle of biologically induced nitrogen removal in wastewater treatment ficilities is based on the activity of populations of autotrophic nitrifying bacteria and their capability to oxidize ammonia (NH_3-N) to NO_2 and NO_3. In addition to nitrification, microorganisms other than the nitrifiers require nitrogen for growth. The amount of nitrogen assimilated during oxidation of carbonaceous material has been placed at 5% of the oxygen demand (C : N : P = 100 : 5 : 1). That means removal of NH_3-N during biological treatment of wastewater may be because of assimilation, not necessarily nitrification. A production of NO_2 + NO_3 as in Figure 6, however, indicates nitrification.

As mentioned earlier, nitrification is affected primarily by pH, DO and temperature. Also, at neutral pH levels there is usually insignificant nitrification until soluble BOD has been oxidized (5). Hence, to evaluate the RBC performance for NH_3 removal, progression of treatment within the RBC stages must be assesed. Oxidation of carbonaceous substrate, expressed as soluble COD is presented in Figure 10 and Table III. The observed COD decrease coupled with the NO_2 + NO_3 increase shown in Figure 6, clearly indicate that the conditions were amenable for nitrification. The impending nitrification was also suggested by the decrease in alkalinity and soluble Total Kjeldahl Nitrogen (TKN) in Table III.

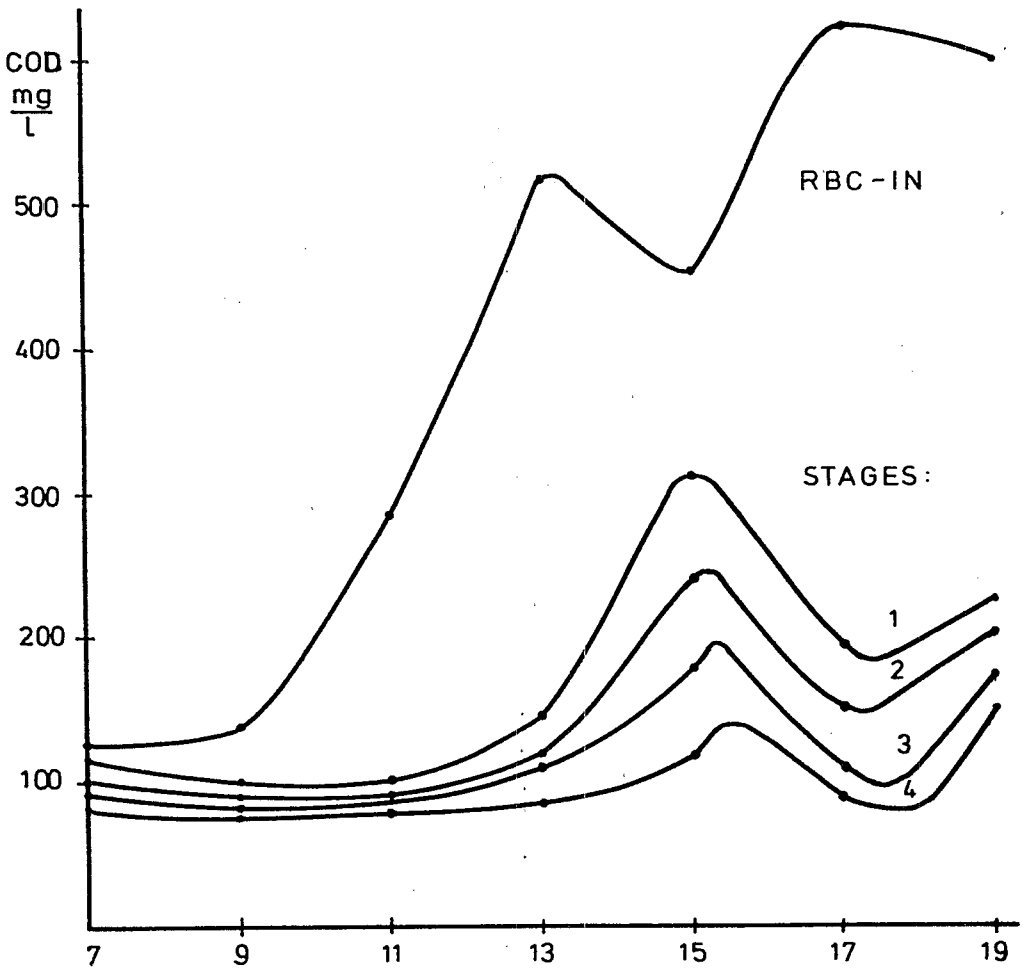

Figure 10. Reduction of COD through the RBC for a typical slaughterhouse production day.

There is no doubt that the BOD removed by the bio-tower was the more readily biodegradable fraction of the wastewater. The soluble COD in the RBC-effluent was approximately 100 mg/l and probably would not decrease much more if further biological treatment was provided. This study supports the evidence that when wastewater is sent through a series of trickling filters, or recycled several times through the same filter, the removability of wastewater organics decreases as the number of passes increases (6).

CONCLUSIONS

The application and start-up performance of a RBC pilot plant unit for upgrading clarified trickling filter effluent has been described. Although the operational experience has been very short, the following remarks can be made from treatment of a 10°C slaughterhouse wastewater:

1. RBC process stability with respect to BOD removal was reached in approximately three weeks after start-up. The RBC-biofilm became mature relatively fast due to seeding of microorganisms from the bio-tower.
2. The hydraulic loading to the RBC was constant at 0.05 $m^3/m^2 \cdot d$ whereas the organic loading varied from approximately 2 g soluble $BOD_7/m^2 \cdot d$ to 35 g soluble $BOD_7/m^2 \cdot d$. A typical slaughterhouse production day organic loading is approximately 20 g soluble $BOD_7/m^2 \cdot d$, resulting in a soluble carbonaceous BOD_7 effluent concentration of approximately 35 mg/l.
3. The Sapromat analysis for soluble BOD did not require seeding or dilution of the wastewater to be tested. The nitrifiers present will therefore perform immediately if other environmental conditions are satisfactory.
4. The effect of low temperature wastewater on NH_3-removal could not be verified in the short time after start-up. This will be a question to answer after prolonged RBC operation.

ACKNOWLEDGEMENTS

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EVALUATION OF AN ANAEROBIC ROTATING BIOLOGICAL CONTACTOR SYSTEM FOR TREATMENT OF A MUNITION WASTEWATER CONTAINING ORGANIC AND INORGANIC NITRATES

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INTRODUCTION

Radford Army Ammunition Plant (RAAP), like most of the Army propellant and explosive manufacturing plants, was built in the early 1940s to supply munitions for World War II. In 1970, the Army initiated modernization programs at its ammunition plants to replace obsolete facilities and improve the safety of operations. As part of this modernization program a continuous automated multi-base line (CAMBL) manufacturing facility was planned for construction at RAAP to augment the present labor intensive batch process. This paper describes the studies that were conducted to develop design criteria for a facility to treat the wastewaters that will be generated in the CAMBL.

WASTEWATER CHARACTERIZATION

A wastewater characterization study was conducted for the CAMBL manufacturing facilities. Samples of the wastewaters were collected and analyzed during the evaluation of a prototype CAMBL manufacturing line. These data were compiled, and the expected characterization of the full-scale facilities were determined. The quantity of water requiring

treatment from the CAMBL facility was determined to be approximately 50,000 gallons per day. This wastewater will contain acetone, ethanol, nitroglycerin (NG), nitroguanidine (NGu), other propellant ingredients, and inorganic nitrates.

LABORATORY TREATMENT STUDIES.

Laboratory-scale treatment studies were conducted to determine the feasibility of selected treatment methods and to define the design parameters for pilot plant studies.

Laboratory studies were conducted to determine the biodegradability of NGu using a biochemical oxygen demand (BOD) test kit. The tests were set up to determine the oxygen uptake rate of the readily biodegradable organic solvents alone, and then with various quantities of NGu added. This testing showed that NGu is not biodegradable by itself, but when combined with a readily biodegradable carbon source NGu is biodegradable.

Studies conducted by Wendt (1) showed that NG is biodegradable, but it does exert a toxic effect on the biological metabolism.

The biodegradability of NG and NGu was further studied using a laboratory-scale rotating biological contactor (bio disc) unit. The wastewater utilized for this study was a mixture of wastewater from the manufacture of other propellants, waste process water from the manufacture of NG, and the CAMBL pilot line effluent. During the study, the bio disc influent contained a chemical oxygen demand (COD) concentration ranging from 500 to 1000 mg/1, a NGu concentration varying from 30 to 70 mg/1, and a NG concentration of approximately 5 mg/1. During this period, the COD removal was approximately 90 percent; the NGu removal ranged between 50 and 90 percent while achieving 100 percent NG removal.

Based upon the wastewater characterization and laboratory studies, two design concepts were considered for this proposed wastewater treatment facility: (a) design a completely new chemical-physical treatment facility for the treatment of this wastewater alone, or (b) expand on the aerobic rotating biological contactor (RBC) treatment plant under construction at RAAP for the treatment of the wastewater from the existing manufacturing facilities.

Alternative (b) was selected for the pilot plant evaluation, based on the estimated savings of over $800,000 in capital costs and an annual savings of about $160,000 in operating costs. The characterization of the wastewaters from the existing manufacturing facilities, the proposed

CAMBL facility, and the combined facilities are shown in Table I.

Table I. Characterization of Waste Waters of Existing and Proposed Facilities

Parameter	Existing Facility	Continuous Facility	Combined Facilities	Increase Due to Continuous Facility
Flow (mgd)	1.245	0.058	1.303	4.6%
COD (lb/day)	7818	1886	6604	40%
(mg/l)			607	
BOD (lb/day)	1887	754	2641	40%
(mg/l)			243	
NO_3 (lb/day)	3024	144	3168	4.7%
(mg/l)			304	
NG (lb/day)	--	14.3	14.3	
(mg/l)			1.50	
NGu (lb/day)	--	57.4	57.4	
(mg/l)			5.28	

The permit issued by the EPA and Commonwealth of Virginia for the wastewater discharge from the aerobic RBC treatment plant was based upon this facility treating the wastewater from the present manufacturing facilities only. This required that any new manufacturing facility to be constructed at RAAP must also provide facilities for treatment of the wastewater generated by that facility to ensure the effluent quality is not degraded.

The aerobic RBC plant hydraulic capacity, as designed, will be adequate for the additional wastewater flow, but additional facilities will be required for the removal of the additional organics, NG, NGu, and inorganic nitrates. The laboratory studies showed that an aerobic biological treatment system appeared to be a suitable method for the removal of the organics, NG and NGu. However, an alternate treatment

method will be required for the removal of the inorganic nitrates. This can best be accomplished by a biological denitrification system. Since additional RBC units will be required for the organic removal, the decision was made to evaluate on a pilot plant scale the use of submerged RBC units for the biodenitrification process.

PILOT PLANT EVALUATION

For the evaluation studies to develop the design criteria for the treatment of the combined RAAP wastewaters, a one-half meter bio surf pilot plant, capable of independent operation of each stage as aerobic or anaerobic, was purchased from Autotrol Inc. This bio surf pilot plant consisted of a series of 36 corrugated polyethylene discs containing a total of 250 ft^2 of surface area. The discs and tank were divided into four stages, separated by removable bulkheads. Each bulkhead consisted of a top and bottom section, whereby, each stage could be operated either completely or 40 percent submerged.

The first phase of the pilot plant evaluation was conducted with all four stages completely submerged to determine the feasibility of the decomposition of the organic solvents, NG and NGu, under anaerobic conditions and to determine the rate of nitrate reduction in a biological denitrification system. An airtight cover was installed on the pilot plant to prevent the diffusion of oxygen into the wastewater from the atmosphere. The system was operated during this period at a hydraulic loading of 1.6 gpd/ft^2 of surface area, and average organic loadings of 6.0 pounds COD and 2.4 pounds BOD per day per 1000 ft^2 of surface area. The NG and NGu concentrations were both maintained between 1 and 5 mg/1. During this phase of the evaluation, the unit averaged 84 percent COD, 90 percent BOD, 94 percent NGu, and 100 percent NG removal. The nitrate removal rate was calculated as a ratio of the BOD removal. The BOD/NO_3 removal ratio during this phase of the evaluation was 0.39. This evaluation demonstrated the feasibility of treating this wastewater by a biological denitrification process to achieve the proposed discharge limitations.

During the second phase of the evaluation, the third and fourth stages of the pilot plant were converted to aerobic stages. This study was conducted to determine the effects of anaerobic RBC units operating in series with the aerobic RBC units. This change had little or no effect on the organics, NG and NGu, removal rates of the pilot plant. Figure 1 shows the COD influent and effluent concentration for this phase of the study.

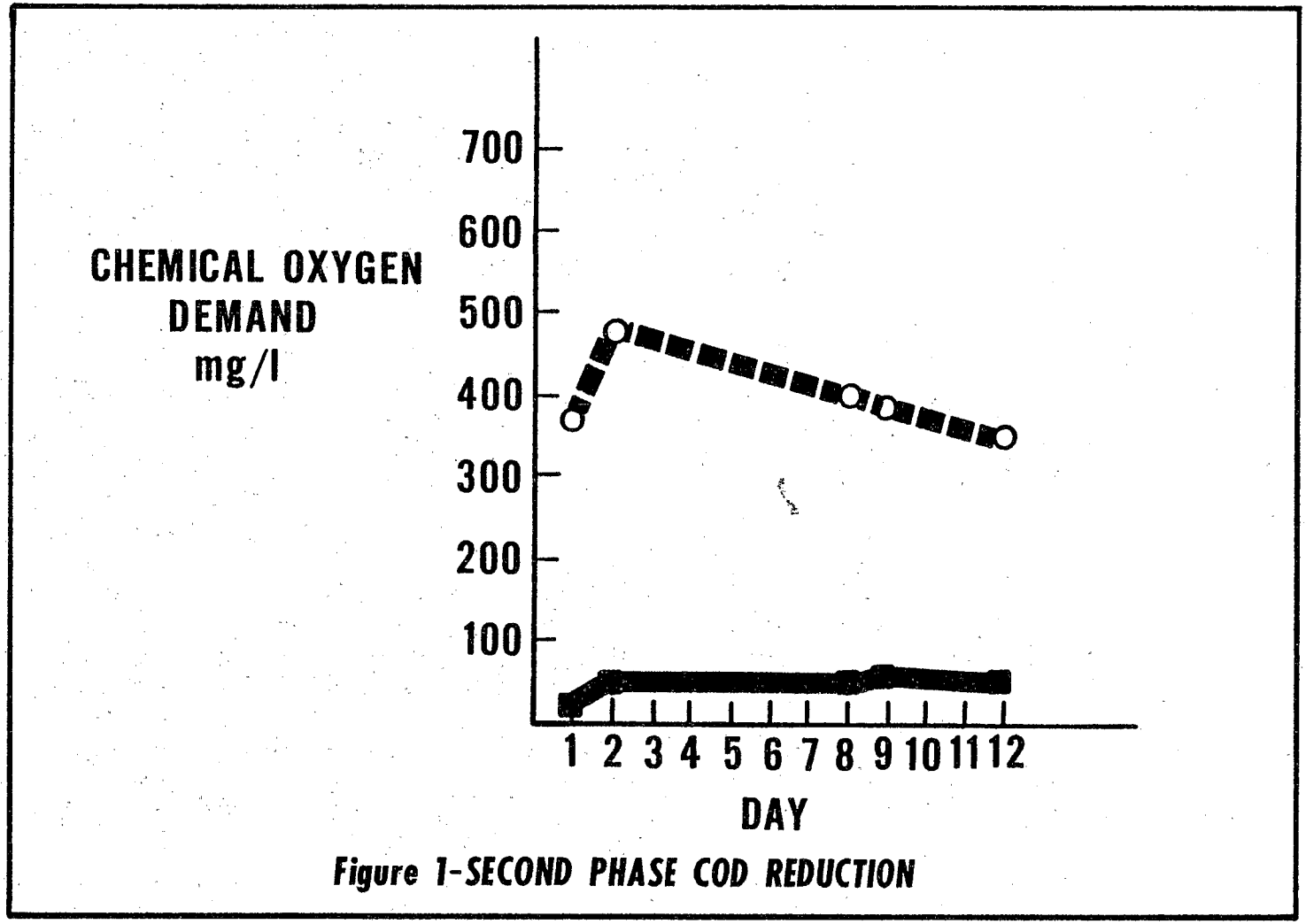

Figure 1-SECOND PHASE COD REDUCTION

The laboratory and preliminary pilot plant data indicated that a biological denitrification RBC system followed by aerobic RBC units is a feasible treatment method for the removal of the organics, NG, NGu, and inorganic nitrates from the CAMBL manufacturing facility. A preliminary design of a system for the treatment of the combined wastewaters consisted of four additional completely submerged RBC shafts preceeding the eight aerobic RBC shafts under construction. To evaluate the efficiency of this proposed system, the pilot plant bio surf unit was converted to a four-stage system, the first stage anerobic followed by three aerobic stages. The sample collection points for the evaluation were selected at the first stage influent and effluent, and the third stage effluent; therefore, simulating the results from the proposed full-scale facility.

During the first two weeks of this evaluation the COD influent concentration was maintained between 400 and 600 mg/1, the NG concentration approximately 1 to 5 mg/1, and the NGu concentration between 10 and 20 mg/1. Figure 2 shows the BOD and COD influent and third stage effluent concentration during this period. The desired influent COD during this period was 450 mg/1. However, due to the constant mixing of wastewaters and the volatility of the organic solvents, wide day-to-day fluctuations occurred. It can be seen from Figure 2 that even with these influent fluctuations, the effluent remained quite constant.

Studies were conducted during the last month of the evaluation to determine if this RBC system could operate effectively under the worst conditions expected in a full-scale facility and still produce an effluent meeting the required discharge standards. The system was operated at an average organic loading of 1.3 times the design loading, NG loading of 3.5 times the design loading, and an NGu loading of twice the design loading. The system was operated at a low temperature of from 6 to 12°C during this period. See Figure 3 for the results of this evaluation. Figure 4 shows the average BOD and COD remaining and the cumulative BOD and COD removal efficiency after each stage of treatment for this study. During this phase, the allowable daily average COD of 190 mg/1 was exceeded on only two days; however, the maximum daily COD effluent concentration of 290 mg/1 was never exceeded. These adverse operating conditions reduced the average COD removal from 85 percent to 74 percent during this period. The NG and NGu removals during this period were near 100 percent most of the time.

The data from the pilot plant bio surf evaluation were analyzed to determine the ratio of organic removal rates to

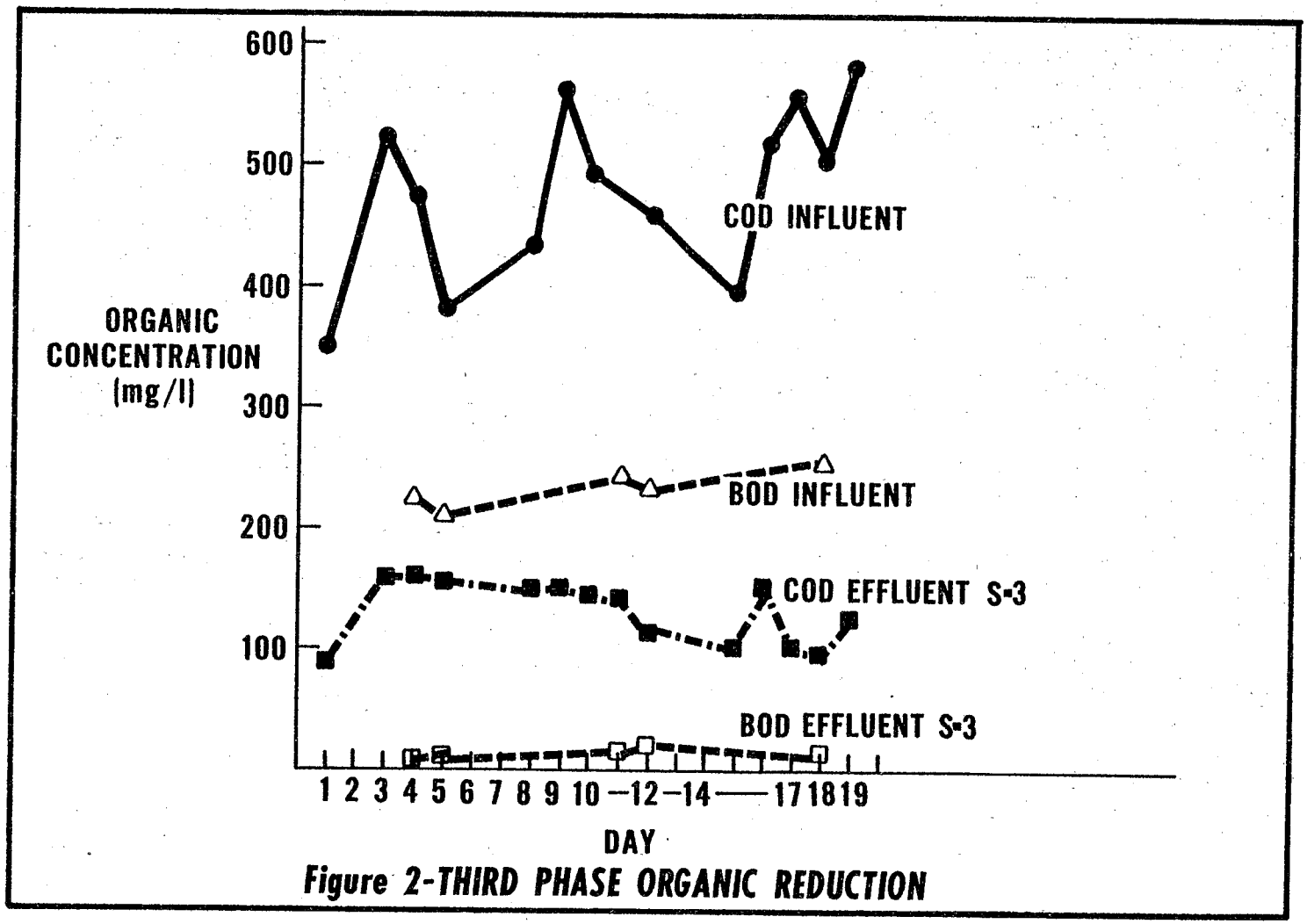

Figure 2-THIRD PHASE ORGANIC REDUCTION

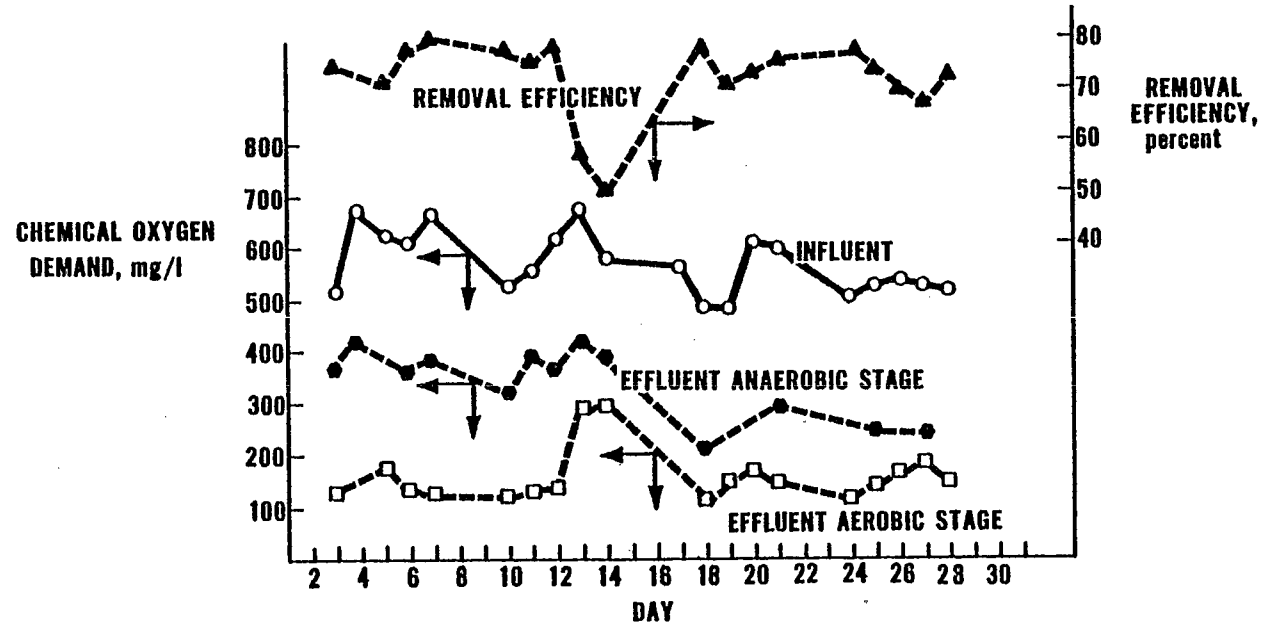

Figure 3-FOURTH PHASE ORGANIC REDUCTION

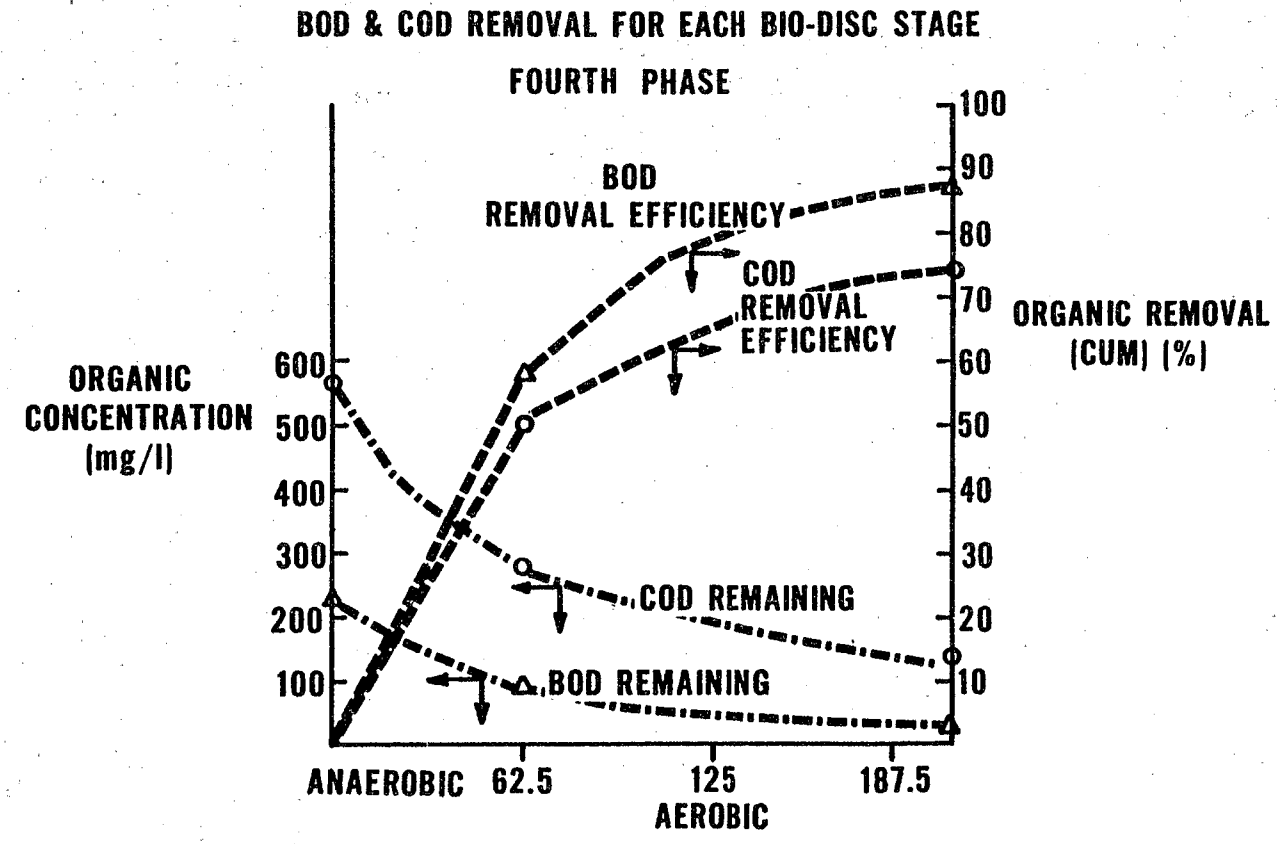

Figure 4-BIO-SURF PILOT PLANT OPERATION AVERAGE CUMULATIVE

the inorganic nitrates removal rates under various operating conditions. As stated above, during the first phase of the evaluation the pilot plant cover was installed to provide a completely anaerobic system. This evaluation showed the BOD/NO_3 removal ratio to be 0.39, which would calculate to provide an average nitrate removal rate of 13.6 lb/day/1000 ft^2 for the full-scale system. During the later evaluations, the bio surf cover was removed and the wastewater in the anaerobic stage was exposed to the atmosphere, allowing oxygen to diffuse into the wastewater, greatly reducing the nitrate removal rate (see Figure 5). Based on the results of this phase, the nitrate removal rate was calculated to be 3.4 lb NO_3/day/1000 ft^2 for the full-scale system. The great differences in the nitrate removal rate between a covered and uncovered system can provide a method to control the nitrate utilization of the submerged RBC stage in the full-scale facility. The system can be designed with removable cover section to provide a flexibility to compensate for low or high nitrate concentrations in the facilities influent.

The results from these evaluations were analyzed to provide the data necessary for the preparation of the design criteria for the proposed facility. The organic load applied was plotted versus the organic load removed for the pilot plant anaerobic system (see Figure 6). A similar graph was also prepared for the first two aerobic stages (Figure 7). These graphs can be used to predict the efficiencies of the full-scale facility at various organic and hydraulic loading. The design of the expanded system at RAAP was analyzed, using these graphs.

Based upon the data from this evaluation, the design criteria for an addition to the RAAP RBC treatment facility were prepared. Figure 8 shows the flow diagram of the proposed RAAP facility for the treatment of combined wastewater. These design criteria were submitted to the Corps of Engineers for the design of the addition to the facility. The final design was completed and construction of the facility at RAAP has been initiated.

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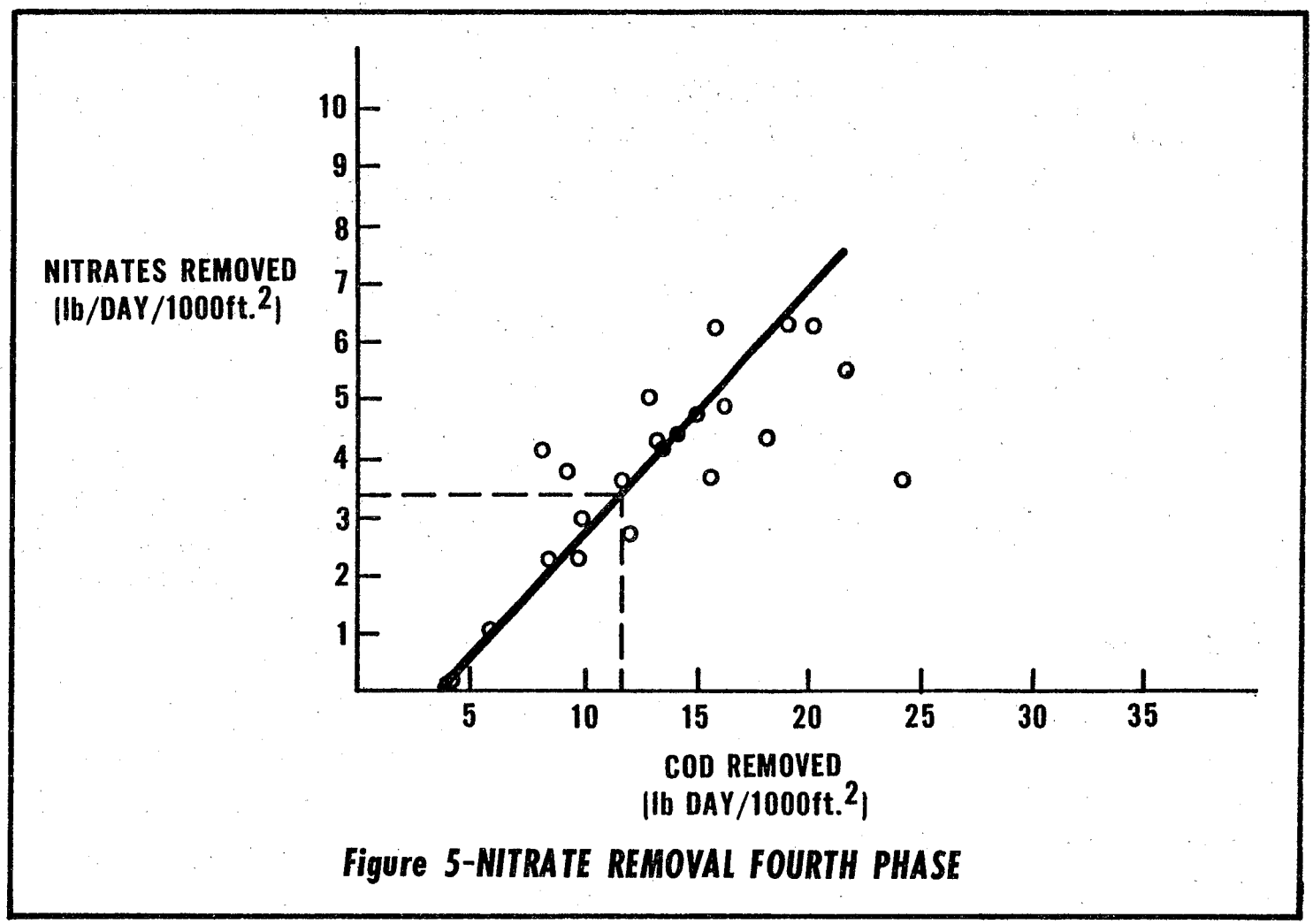

Figure 5-NITRATE REMOVAL FOURTH PHASE

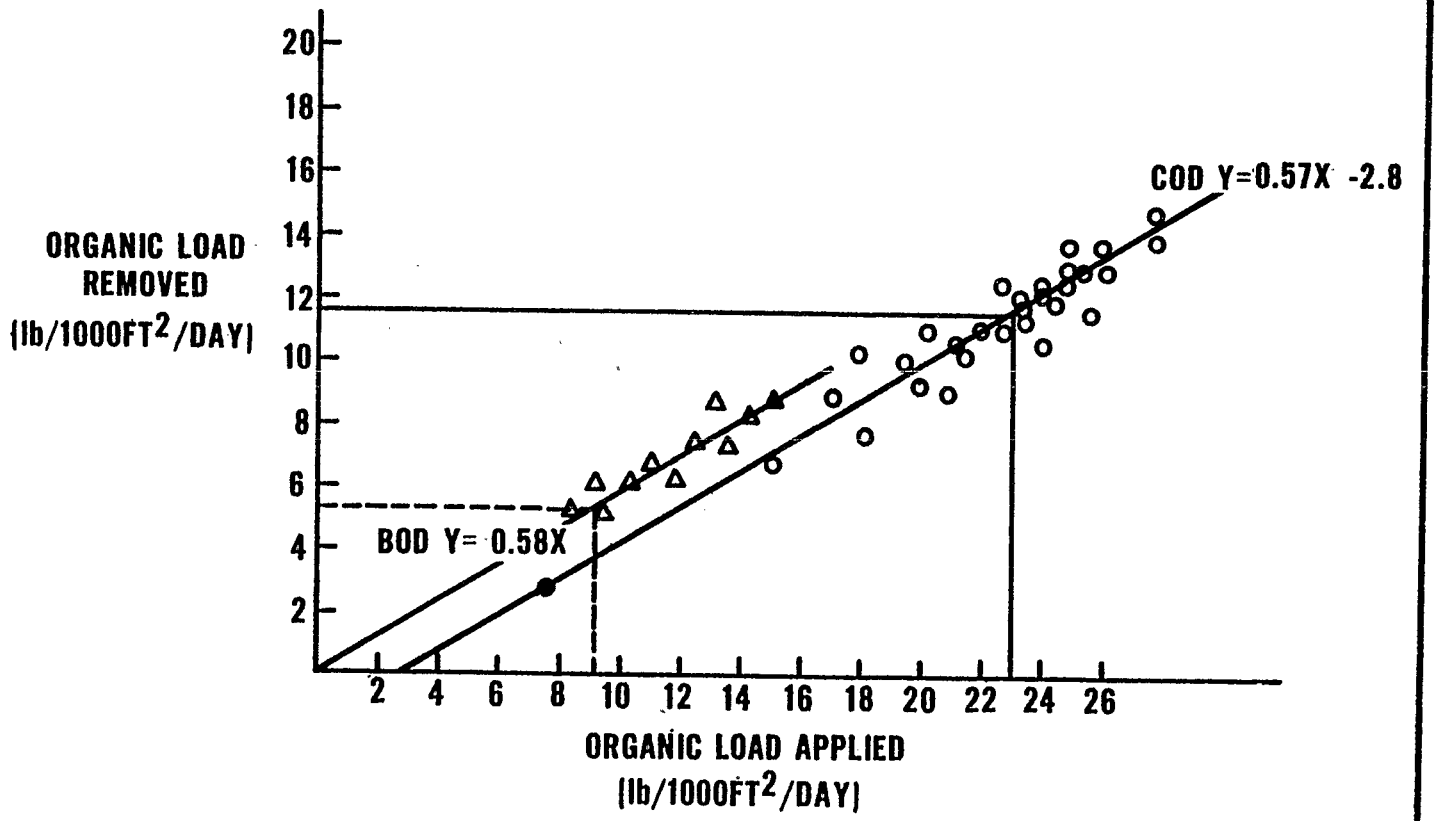

Figure 6-ORGANIC LOAD REMOVAL ANAEROBIC STAGE

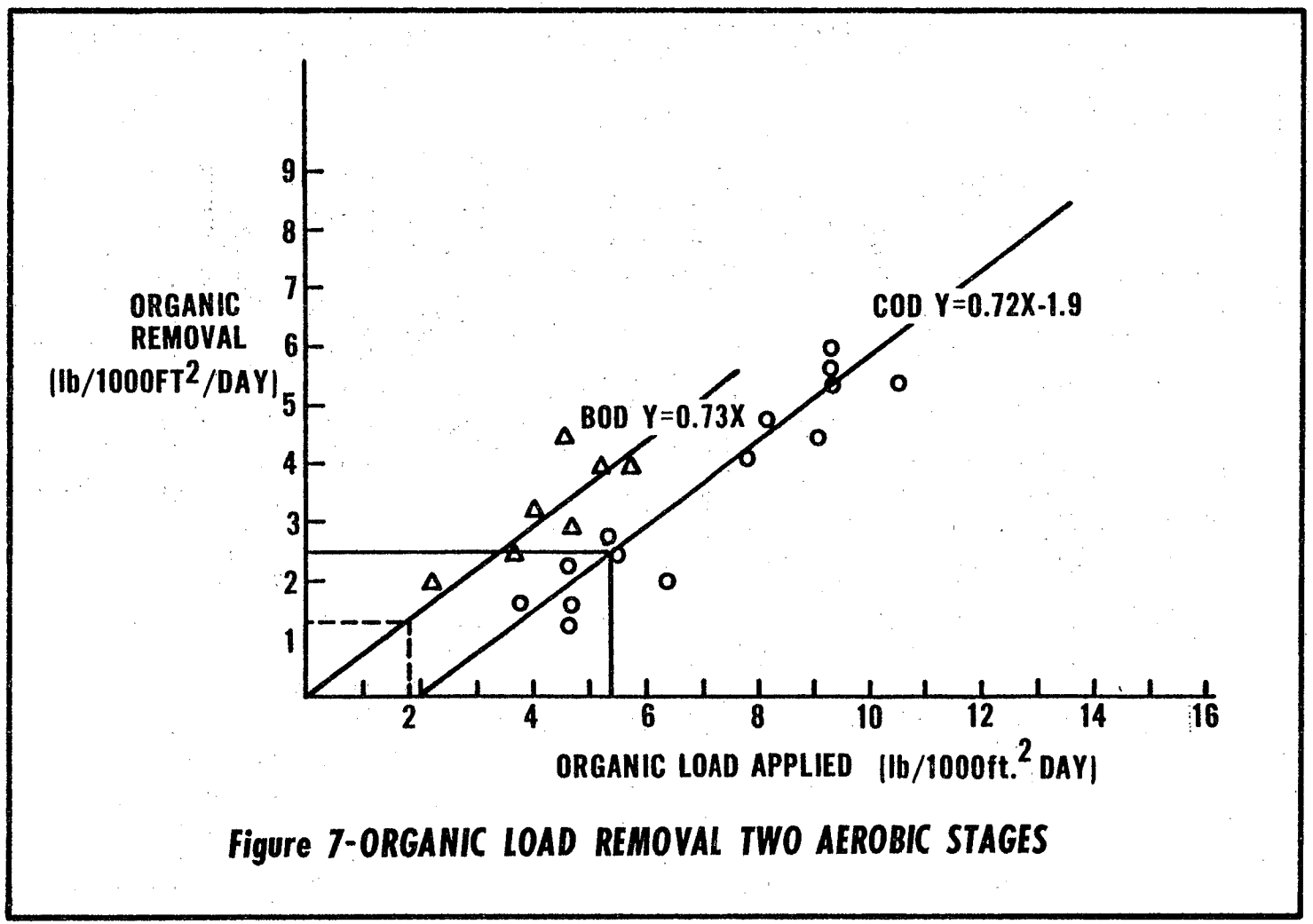

Figure 7-ORGANIC LOAD REMOVAL TWO AEROBIC STAGES

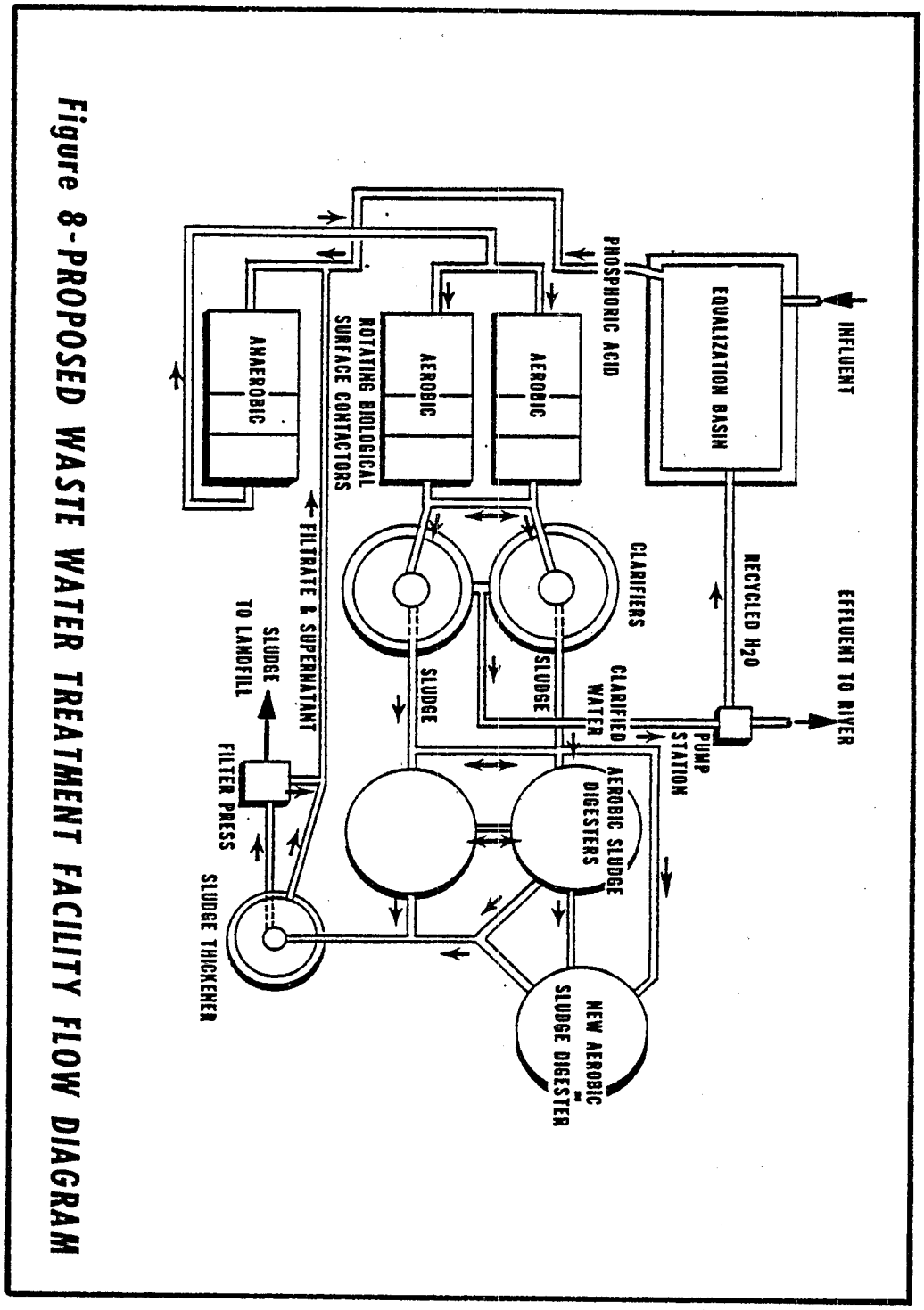

Figure 8-PROPOSED WASTE WATER TREATMENT FACILITY FLOW DIAGRAM

APPLICATION OF ROTATING BIOLOGICAL CONTACTOR (RBC) PROCESS FOR TREATMENT OF WASTEWATER CONTAINING A FIREFIGHTING AGENT (AFFF)

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INTRODUCTION

A firefighting agent, Aqueous Film Forming Foam (AFFF), has been used for fuel/oil fire extinguishment at airports and on shipboard since 1970's. AFFF has been found to be the most effective fuel/oil firefighting agent ever to be formulated. In accordance with firefighting performance specifications used by the manufacturers, a 28-square foot fuel fire can be extinguished within 45 seconds with a 6% AFFF solution (by volume).

AFFF consists of fluorochemical surfactants, hydrocarbon surfactants, ethylene glycol and its derivatives, and about 70% water (by weight). In a firefighting operation, the AFFF concentrate is diluted to a 3-6% solution (by volume), and sprayed under pressure onto the fire. The foam created during the spray covers and extinguishes the fire. AFFF concentrate (FC-780) contains an organic load of approximately 380,000 mg/L COD, or 110,000 mg/L TOC or 325,000 mg/L BOD (Ref 1). A toxicity test with fathead minnows indicated that the 48-hour TLm (LC_{50}) concentration was about 1800 ppm (FC-206, by volume) (Ref 2). A maximum

loading rate to an activated sludge treatment process (with acclimation) was found to be 250 ppm (FC-206 by volume) without the addition of an antifoam agent (Ref 2).

The Rotating Biological Contactor (RBC) is considered to be a most cost-effective wastewater treatment process due to it being simple in operation, low in capital investment and low in energy requirements. The RBC will fit well with trickling filter systems, which constitute approximately 95% of the sewage treatment systems on military bases, and will upgrade the effluent water quality to meet the National Pollutant Discharge Elimination System (NPDES) permit standards.

Increased popularity in the use of the RBC on military bases is anticipated. This, in part, is due to the successes experienced by other researchers for using the RBC to treat various organic compounds, such as formaldehyde and formic acid as well as the explosives RDX, HMX and TNT (Ref 3). Such research prompted the following experimentation for determining the RBC's feasibility (technically and economically) for treating AFFF containing wastewater. The research effort initially began with a chemostat study of the parameters and microorganisms that were amenable to AFFF bioconversion. This was followed by experimentation with a four-stage bench top RBC system. The percent (%) conversion in COD, BOD and TOC was monitored as a means of determining removal of AFFF.

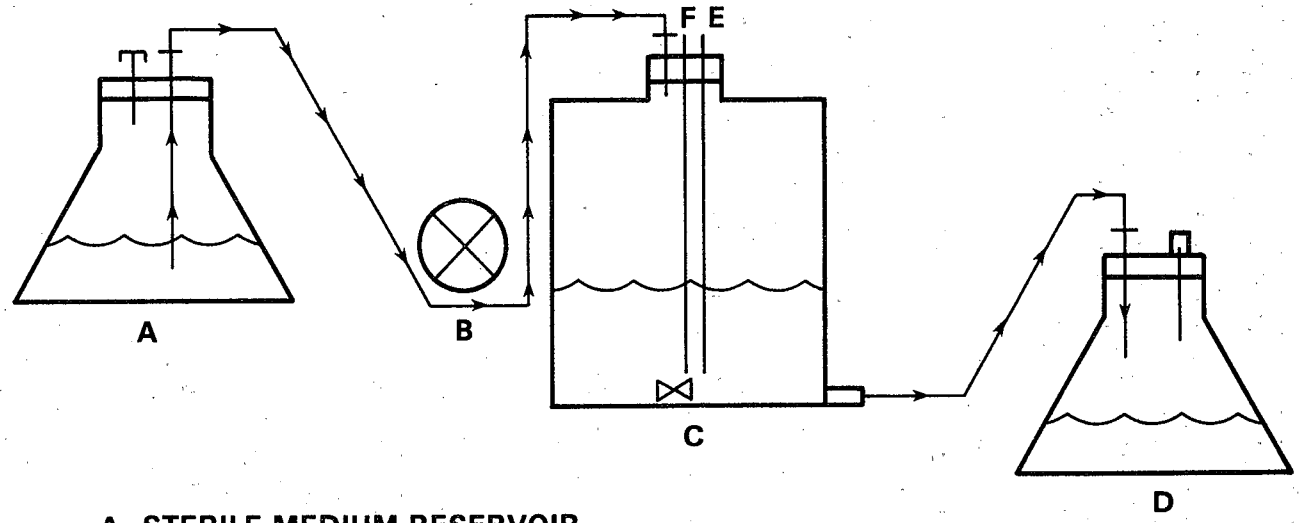
MATERIALS AND METHODS

Two types of experimental systems were used in this study. One was an aerobic chemostat used as an approach to determine initial feasibility, the other was a bench top model of an RBC.

Chemostat

Physical Set-Up: A diagram of the chemostat physical set-up is given in Figure 1. Influent was sterilized in a two-liter reservoir and put on line aseptically. The flow rate of sterile medium into the reaction vessel was 1 ml/min. The reaction vessel was a 4-liter aspirator bottle (Kimax), which was continuously agitated via a stir bar/stir motor arrangement (Corning Hot Plate Stirrer, PC-351). Aeration was accomplished via filtered air (Acropore 0.45-µm filter) bubbled into the bottom of the reaction vessel. A

CHEMOSTAT



A - STERILE MEDIUM RESERVOIR
B - PERISTALTIC PUMP
C - REACTION VESSEL
D - WASTE
E - SAMPLE PORT
F - AERATION

Figure 1

constant volume of 2 liters was maintained within the reaction vessel by a siphon overflow tube into a waste receptacle which was replaced and autoclaved when full. Samples were taken by suction-draw from the sampling port.

Inoculum: The inoculum or seed for the start up of the chemostat was 0.05 gm each of the following: dried bacteria culture (Horizon Ecology Company) for degrading fats, oils and greases (#245-40), for hydrocarbon degradation in fresh water (#245-60), and 5 ml of activated sludge from the Buena Ventura County Water Treatment Plant.

Media: Bushnell Haas Broth (Difco) was used as a minimal salts medium to which specific amounts of known carbon could be added. The carbon used in this experiment were D-Glucose (Difco) and/or the aqueous film forming foam designated FC-780 (3M). This then comprised the sterile influent.

Growth conditions: The chemostat experiment was conducted at ambient temperature, under mild aeration and agitation. pH was monitored but no attempt at adjustment was made.

Procedure: The system start-up was as follows. The reaction vessel containing two liters of sterile Bushnell Haas Broth (BHB), 0.05% Glucose and 0.5% FC-780 was seeded with the inoculum and allowed to grow as a batch system. After 48 hours and an increase to 0.60 optical density, sterile influent containing 0.05% Glucose and 0.5% FC-780 was put on line. On day 8, the influent was changed to contain 0.5% FC-780 (approximately 2000 ppm COD) as the only carbon source. Samples from influent & effluent concurrently, were taken three (3) times per week and analyzed as follows:

A. Turbidity. Utilizing the sterile influent as a standard or blank, turbidity of the effluent was determined at 460 nm, utilizing a Beckman Spectronic 88.

B. pH. The pH of the influent and the effluent were determined immediately after sample withdrawal, using an Orion Research Model 701A/Digital Ionalyzer.

C. Biochemical Oxygen Demend (BOD). The 5 day BOD determination was used as outlined in Section 507 in Standard Methods (Ref 4), utilizing an Orion Research Model 701A/Digital ionalyzer and Mode 97-08-00 O_2 electrode.

D. Chemical Oxygen Demand (COD). COD was performed according to the method outlined in Section 508 of Standard Methods (Ref 4) and modified by Technicon (Ref 5).

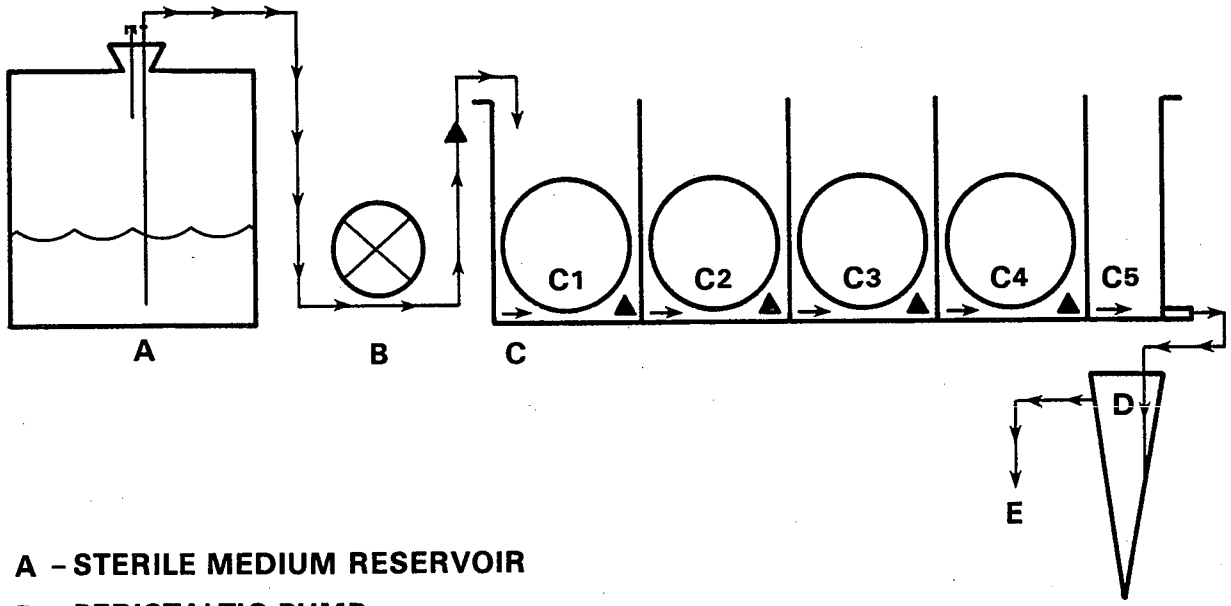
E. Total Organic Carbon (TOC). A variation on the procedure given in Section 505 of Standard Methods (Ref 4) was used. The variation, the acid sparge technique, was performed with the Beckman 915B TOC analyzer, and is outlined in the operation manual (Ref 6). All samples used for TOC, COD and/or BOD determination were filtered prior to analysis through a series of graded membrane filters, i.e. 5 µm, 1.2 µm, 0.8 µm and 0.45 µm (Gelman). Each filter was washed prior to use with 30 ml of double deionized water to remove any organic wetting agent on the filter.

F. Microorganism Identification. Bacterial and fungal populations were identified and enumerated utilizing Nalgene Nutrient Pad Kits, of the following media: Standard TTC - for total counts, Azide - for enterococci and fecal streptococcus, Wort - for fungi, filamentous and non-filamentous, Weman - for slime forming mesophilic bacteria (e.g. Leuconostoc mesenteroides).

Rotating Biological Contactor (RBC)

Physical Set-Up: A diagram of the RBC physical set-up is given in Figure 2. Influent was sterilized and aseptically added to a 20-liter reservoir (5 gal. bottle, Kimax). The reservoir was then put on line aseptically. The flow rate of sterile influent into the aerobic RBC was 3.5 ml/min, and was controlled by a peristaltic pump (Cole Parmer). The RBC was on loan from the U.S. Army Mobility Equipment Research and Development Command (USAMERDC), Ft. Belvoir, VA., and has been described in detail by them (Ref 3). Basically, it was a five chambered unit constructed of Plexiglass. Each of the first four chambers contained six

ROTATING BIOLOGICAL CONTACTOR



A – STERILE MEDIUM RESERVOIR

B – PERISTALTIC PUMP

C – R B C

D – CLARIFIER

E – WASTE

▲ – SAMPLING POINTS

Figure 2

1/4-inch (0.6 cm) thick plexiglass discs, 9-1/2 inches (24 cm) in diameter mounted on a shaft 1/2 inch (1.27 cm) in diameter. One hundred and twelve (112) holes, 1/4 inch (0.6 cm) in diameter, were bored into each disc to aid in microbial attachment. The total disc area was 23.55 ft^2 (2.188 m^2). The last chamber was void of discs, acting as a 1 liter capacity reservoir-clarifier. The total liquid capacity of the unit was 14.5 liters. The discs were rotated at 17.5 rpm, thus being equivalent to an edge velocity of 0.73 ft/s (0.22 m/s). An additional clarifier was added in the form of a modified Imhoff cone, which was used to visually measure the amount of sedimentation produced in a 24 hour period. Samples were taken from within all four stages and from the influent.

Inoculum: The seed for the start-up of the RBC was one liter of activated sludge obtained from the Buena Ventura County Water Treatment Plant, and was inoculated within one hour of acquisition.

Media: BHB was used. Varying concentrations of FC-780, D-glucose, and Nutrient Broth (Difco) were added as outlined in the procedure.

Growth Conditions: The RBC experiment was conducted at ambient temperature. Aeration was accomplished by the revolution of the discs through the wastewater. pH was monitored and adjustments were made, using 1N NaOH or 1N HCL, when necessary.

Procedure: The system start-up was as follows. The RBC was filled with 14.5 liters of BHB plus 0.1% glucose, inoculated with activated sludge and allowed to run as a static system for 24 hours. Sterile influent containing 0.1% glucose was fed into the unit at a rate of 3.5 ml/min. After 2 days it was determined that this mode of addition of the carbon source was inadequate to maximize colonization of the discs and so glucose and/or nutrient broth was added to each stage once daily to a total concentration of 0.1% carbon. On day 29, FC-780 was added to the influent at a concentration of 100 ppm in terms of COD. The concentration of FC-780 was gradually increased until a level of 1000 ppm COD was achieved. Samples were taken three times per week. BOD, COD, TOC and pH analysis were performed as described under the chemostat procedure. Other parameters measured were:

A. Temperature. Readings were taken three times per week, utilizing a Wahl digital heat-prober thermometer. The thermometer was placed directly into each of the four stages of the RBC.

B. Microorganism Identification. Bacterial and fungal populations were identified and enumerated utilizing: Nalgene Nutrient Pad Kits -- TTC and Wort, Bio Stix and Myco Stix test strips (Ames Company) and/or Total Count Water Tester (Millipore Corp). Microscopic qualitative observations were done every 14 days to visually monitor changes in predominant populations, i.e. protozoal, fungal, and nematodal.

RESULTS AND DISCUSSION

Chemostat

After 7 days of continuous operation, an apparent steady state condition was achieved within the chemostat in terms of COD, TOC and BOD conversion or percent (%) removal from the supernate. Approximately 70% COD conversion, 80% BOD conversion and 60% TOC conversion were consistently observed from day 11 onward to day 43 (Figure 3). From day 40 until shut down of the chemostat on day 63, the percent (%) conversion dropped to approximately 45% COD, 50% BOD and 40% TOC. This was in part correlated with a rise in the pH of the reaction vessel to a pH of 7.1 or greater. The microbial populations observed in the chemostat changed drastically with the increase in pH. That is, a greater number of yeast and slime-forming bacteria were noted. No effort was made to readjust the pH of the chemostat and so the percent conversion in all three dropped to a level of 40-50% conversion. It was decided for future experimentation to adjust the pH of the RBC to 7.00.

As seen in Figure 3, the percent (%) conversion values exhibited some variance. This is partially due to technical errors and machine failure. That is dilution and sampling errors were committed during a turnover of technical assistance. Equipment failure would occur and no new influent would enter the reactor vessel for a 12 to 16 hour period. This would result in microbial back contamination from the reactor vessel into the influent reservoir, which would result in an increase in the pH and a decrease in available

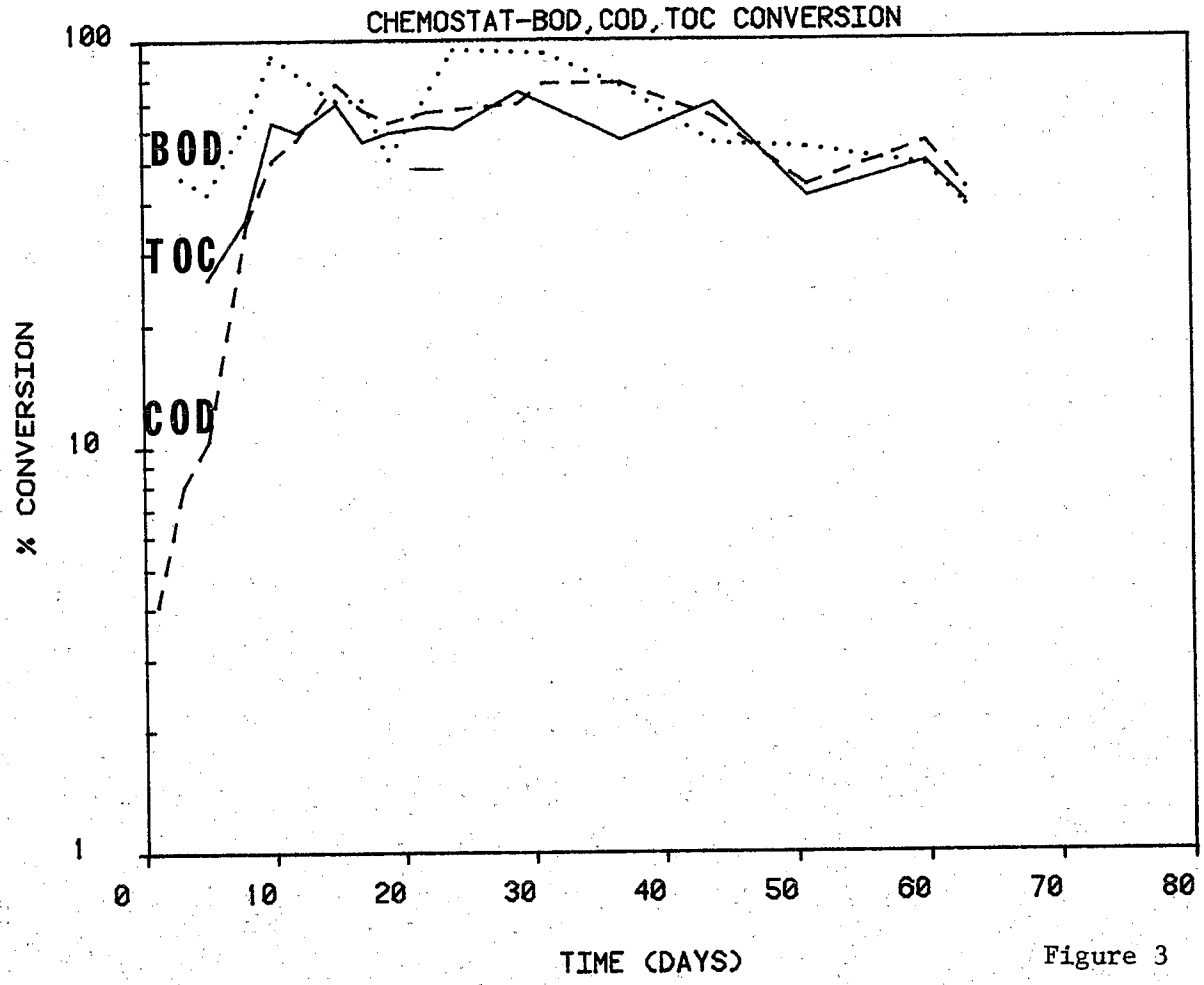

CHEMOSTAT-BOD,COD,TOC CONVERSION
100
10
1
% CONVERSION
BOD
TOC
COD
0
10
20
30
40
50
60
70
80
TIME (DAYS)

Figure 3

carbon in the influent and thus a lower TOC, COD and BOD conversion measurement. It should also be mentioned that plate counts were performed infrequently and were more for qualitative determination of different microbial populations.

From the chemostat data, it was decided that enrichment for mixed microbial populations, that were able to utilize FC-780 as their sole source of supplemented carbon, was possible. The data from the chemostat also indicated that it was possible to change an influent containing 1200 to 1500 ppm COD of FC-780 into an effluent containing 100-200 ppm COD.

Rotating Biological Contactor

At initial start-up, 0.2% glucose was added via the influent to stage 1 of the RBC at a flow rate of 4 ml/min. This proved to be too high of a concentration of glucose in that the pH of the RBC rapidly became acidic and was thought to endanger the not yet well established microbial population. Therefore, the concentration of added carbon, in the form of glucose, was dropped to 0.1%. However, the majority of this carbon (97%) was used in stages 1 and 2, and stages 3 and 4 failed to exhibit growth on the discs. To achieve colonization of all the RBC discs, 0.1% carbon-source, in the form of a 10X concentrate, was added to each stage daily. This also stopped the recurring back contamination into the sterile influent reservoir, which now contained BHB only.

It was noted that the pH of the effluents daily dropped into the acidic range (6.0 - 6.9) and had to be chemically adjusted. After 3 days, nutrient broth was added in the form of a 10X concentrate, along with the 10X glucose, to result in a final concentration of 0.1% carbon. It was thought that whereas glucose was metabolized aerobically into acids, the nutrient broth would be metabolized with the resulting release of amino groups. This would help to raise the pH, and the protein itself would also act as an additional buffer. This provided adequate pH regulation unless a malfunction in the equipment or a laboratory error occurred which resulted in a decrease in the pH of one or glucose was metabolized aerobically into acids, the nutrient broth would be metabolized with the resulting release of amino groups. This would help to raise the pH, and the protein itself would also act as an additional buffer. This provided

adequate pH regulation unless a malfunction in the equipment or a laboratory error occurred which resulted in a decrease in the pH of one or more of the stages.

The resulting reduction in COD, BOD, and TOC are presented in Figures 4, 5, and 6, respectively. In these figures, percent removal is shown with respect to time given in days. After 30 days of continuous operation, an apparent steady state condition was achieved within the RBC, in terms of COD, TOC and BOD removal. Approximately 97% removal was achieved in all three parameters measured. As seen in Figure 7, exposure of the RBC to FC-780 began on day 35 with the addition of 0.025% FC-780, or 100 ppm in terms of COD. By day 60, 1000 ppm COD of FC-780 was being fed. Simultaneously, the amount of nutrient broth, which was the only other carbon source after day 52, was lowered to a level of approximately 500 ppm COD. This level of carbon was maintained until day 80. The conversion rate at that time was 98% COD, 96% BOD, and 94% TOC. A one way completely randomized analysis of variance was conducted on each stage with respect to COD, BOD or TOC. These results are given in Table I.

Table I. RBC, One-Way Completely Randomized Analysis of Variance Versus Bartlett's Variance

Analytical Form	One-Way Analysis F	One-Way Analysis Significance	Bartlett's Variance F	Bartlett's Variance Significance
COD	441.846	0.000	34.543	0.000
BOD	175.475	0.000	17.659	0.000
TOC	53.871	0.000	91.259	0.000

The calculated $F_{.01,4,25}$ value for this test would be 3.47. The values shown in Table I, being larger than the calculated F, are indicative of a significant variation between treatment and non-treatment with the RBC. The Bartlett's test of homogeneous variance indicates no violation of the homogeneity assumptions of ANOVA. The significance levels show low probability of error within the tests.

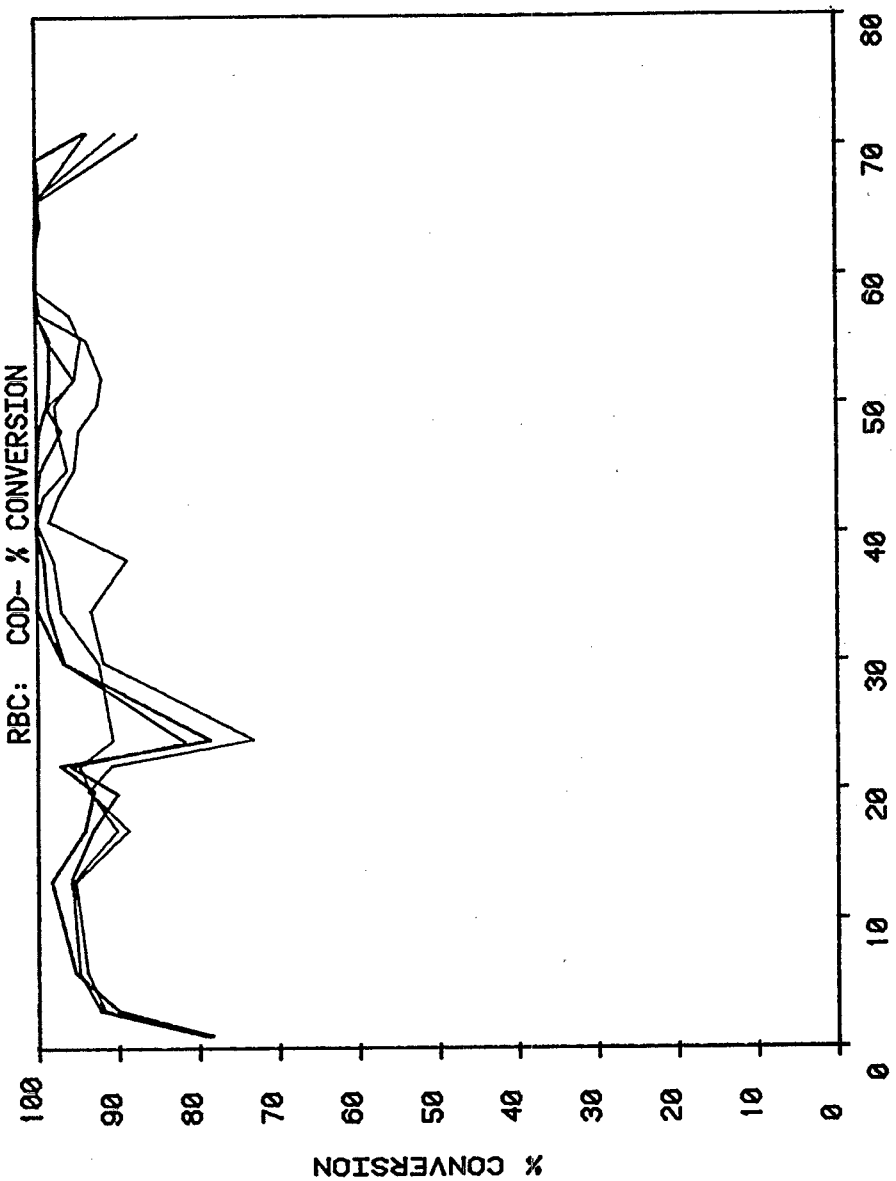

RBC: COD- % CONVERSION
% CONVERSION
100
90
80
70
60
50
40
30
20
10
0
0
10
20
30
40
50
60
70
80
TIME (DAYS)

Figure 4

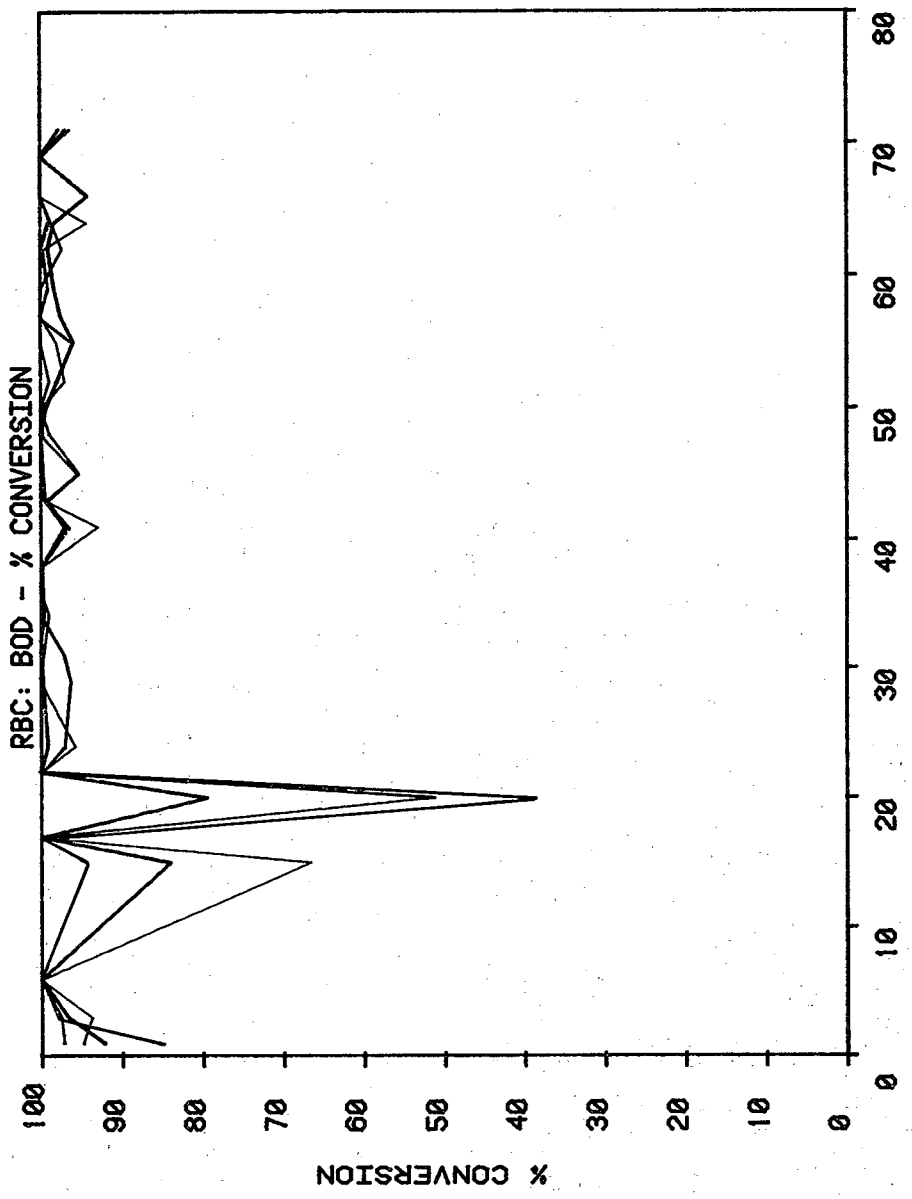

RBC: BOD - % CONVERSION
% CONVERSION
TIME (DAYS)
0
10
20
30
40
50
60
70
80
90
100

Figure 5

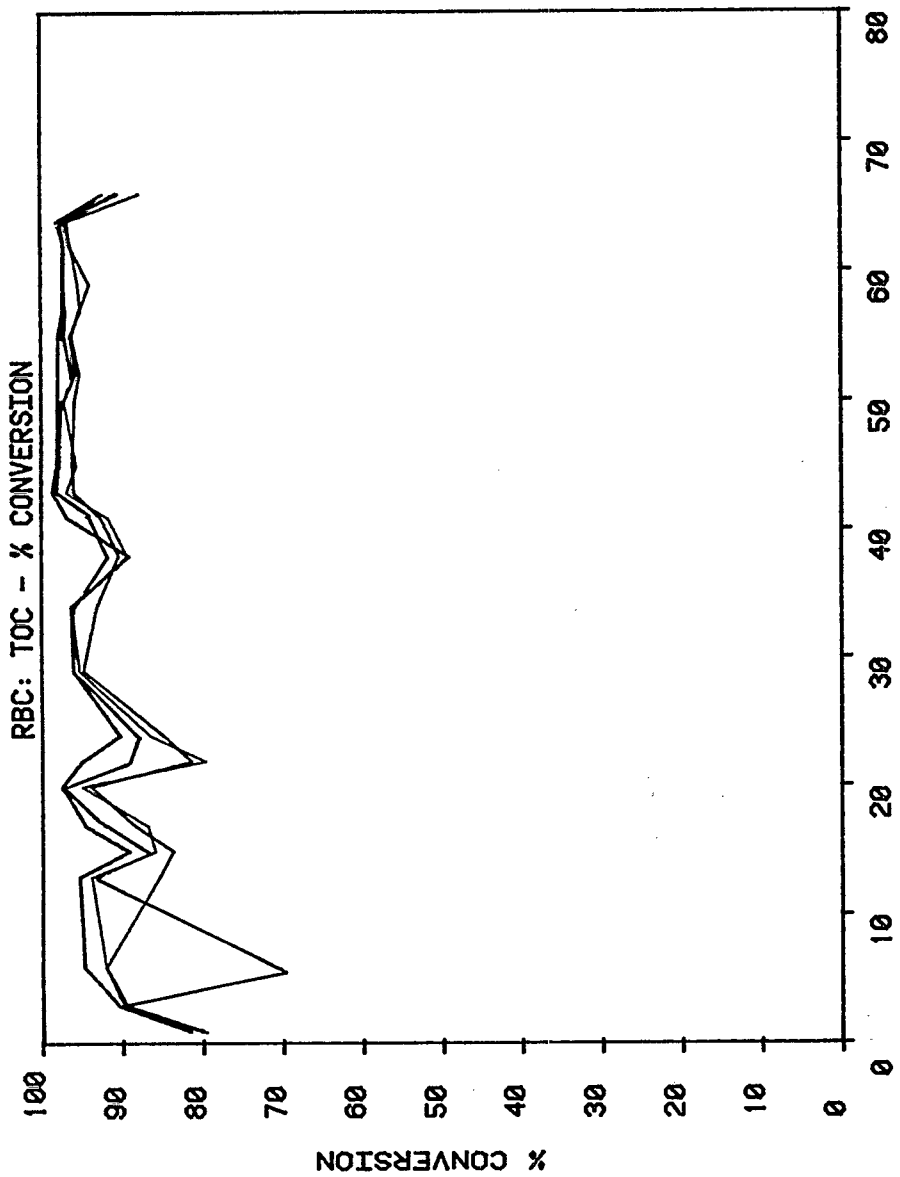

RBC: TOC - % CONVERSION
% CONVERSION
100
90
80
70
60
50
40
30
20
10
0
0
10
20
30
40
50
60
70
80
TIME (DAYS)

Figure 6

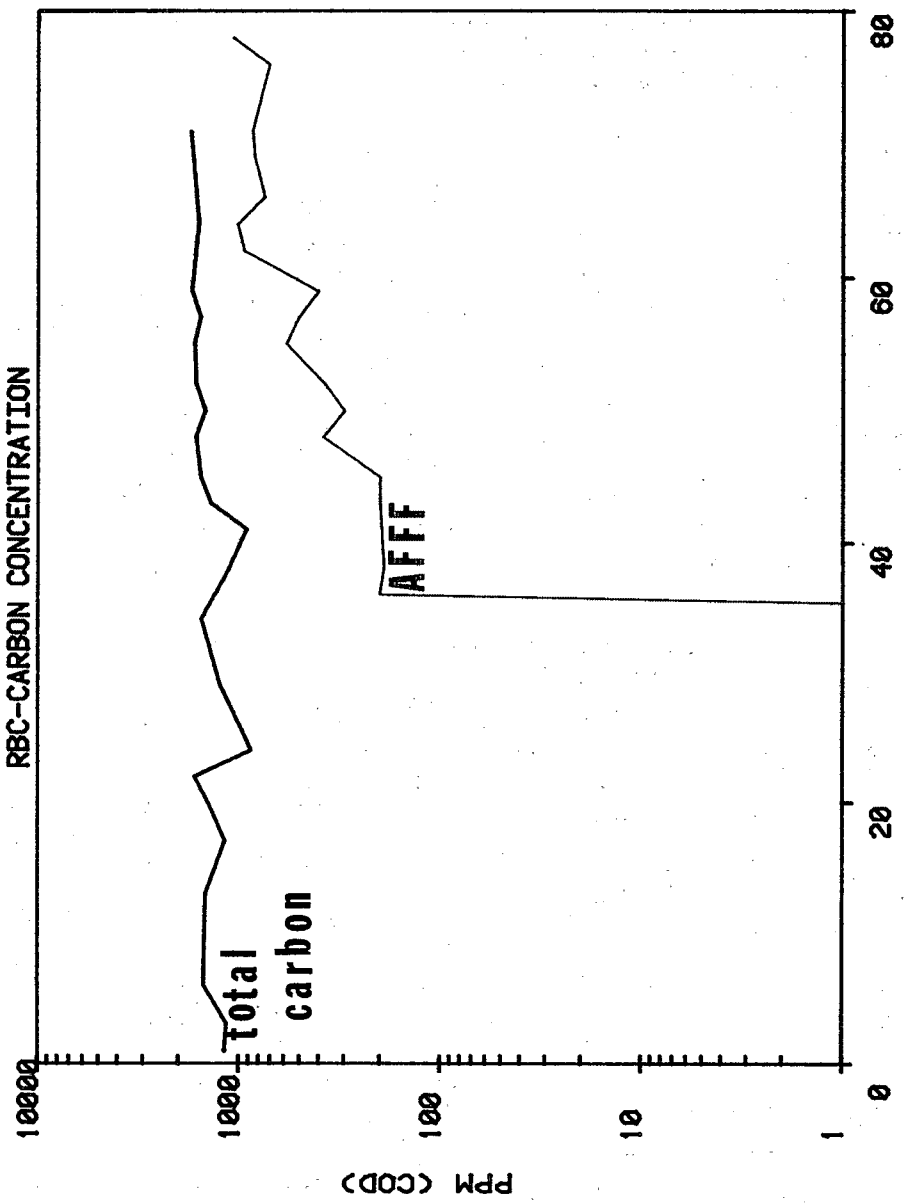

RBC-CARBON CONCENTRATION
10000
1000
100
10
1
PPM (COD)
total carbon
AFFF
0
20
40
60
80
TIME (DAYS)

Figure 7

It can be seen from these results that the use of the RBC was very significant in treating wastewater containing up to 1000 ppm COD of FC-780.

Microbial populations observed are given in Table II. A strong, heterogeneous population was observed throughout the experiment. Although some changes in densities occurred, the organisms listed in Table II were seen throughout the experiment.

CONCLUSION

The purpose of this research was to determine the feasibility of treating wastewater containing aqueous film forming foam (FC-780) by an aerobic RBC. The preliminary data presented here demonstrates that FC-780 is conducive to aerobic bioconversion and removal with a properly adapted RBC unit. A significant reduction in COD, BOD and TOC has been achieved in synthetic wastewater containing up to 1000 ppm COD of FC-780. It is possible that FC-780 loading may be increased further and that higher reduction of the parameters may be obtained by changes in flow rates and contact times. These possibilities are being actively addressed in preparation of scale up for pilot plant operation.

Table II. Microbial Groupings as Observed on Suspended Microscope Slides in the Aerobic Rotating Biological Contactor

Nematodes	Bacteria
Fungi	Gram Positive
Filamentous	Staphylococci
Ex: *Aspergillus*	Streptococci
Pennicillium	Bacillae
Non-filamentous	Gram Negative
Protozoa	Bacillae
Sarcodina	Filamentous bacillae
Ciliata	Algae
Ex: *Suctoria*	Phaeophyta
Zoothamnia	
Voticella	
Paramecium	

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OPERATION OF A RBC FACILITY FOR THE TREATMENT OF MUNITION MANUFACTURING PLANT WASTEWATER

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INTRODUCTION

In 1970, the Army initiated an extensive pollution abatement program at all of its ammunition plants. The treatment of waste process waters from these plants required development of new or modifications of existing technology because of the unique nature of the pollutants in the wastewater. This wastewater contains ether, alcohol, acetone, inorganic nitrates, traces of nitroglycerin (NG), and other propellant ingredients.

The initial wastewater treatment studies[1] were conducted using an activated sludge process. This study demonstrated that the activated sludge process was not a feasible treatment method for the RAAP wastewater due to the high variability in flow and organic concentrations. A successful rotating biological contactor (RBC) pilot plant evaluation[1] was conducted to define the design parameters and develop the design criteria for a full-scale facility.

TREATMENT PLANT DESIGN

A RBC wastewater treatment facility was constructed at RAAP for the treatment of the process wastewater based upon

the design criteria developed from the pilot plant evaluation. This facility consists of a 5110 m^3 (1,350,000 gal) equalization basin and eight RBC shafts containing a total surface area of 56,782 m^2 (612,000 ft^2). Since the equalization basin would probably develop a dispersed biological growth, even without the addition of nutrients, four 15-hp floating aerators were provided to mix the basin and prevent sedimentation of suspended solids, and to provide adequate aeration to satisfy the oxygen uptake rate of the dispersed growth.

The RBC system (figure 1) was constructed to provide two separate parallel RBC systems, each system consisting of three stages. Stage one of each system contains two RBC shafts while the other stages contain one shaft in each stage. The design parameters for the RBC facility are shown in table I.

Table I. RBC Design Parameters

	Flow	Chemical Oxygen Demand (COD)	Biochemical Oxygen Demand (BOD)
Avg Flow Rate - m^3/day	4716		
(gpd)	(1,250,000)		
Design Load - kg/day		2140	856
- (lb/day)		(4718)	(1888)
- mg/l		452	181
Avg Hydraulic Loading			
- $m^3/m^2.d$	0.08		
- (gpd/ft^2)	(2)		
Avg Organic Loading			
- kg/1000 $m^2.d$		37.7	15.1
- (lb/day/1000 ft^2)		(7.7)	(3.1)
Discharge Limitations, maximum			
- Average daily (mg/l)		195	60
- Maximum daily (mg/l)		290	120

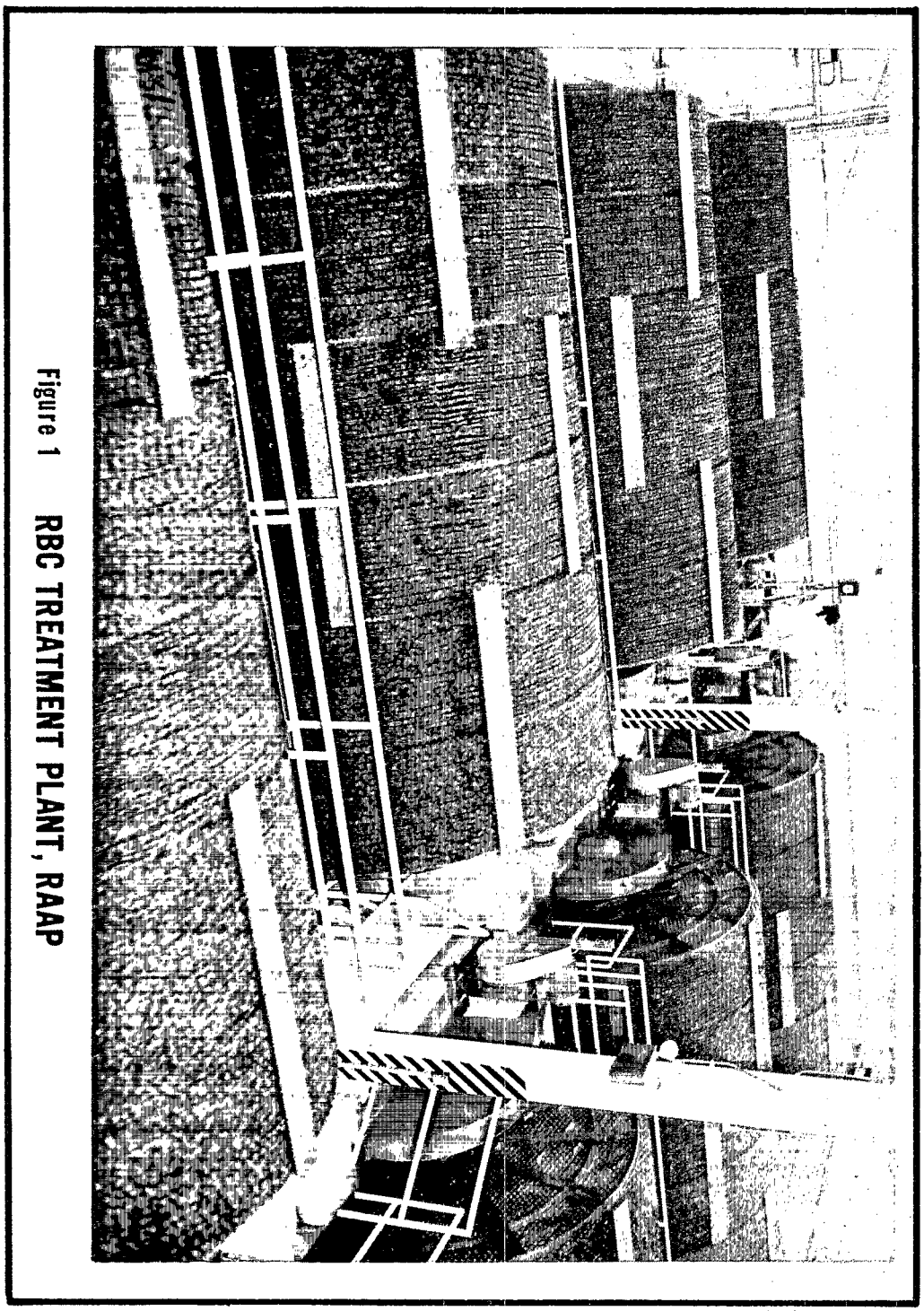

Figure 1 RBC TREATMENT PLANT, RAAP

FACILITY START-UP

The facility was placed into operation during December 1980. The RBC basins were filled with process wastewater from the equalization basin, 3.8 m^3 (1000 gallons) of waste sludge from a local municipal activated sludge wastewater treatment plant, and one liter of phosphoric acid. The RBC system was operated in a batch mode for eight days to allow a biomass growth to develop on the RBC media. During this time ethyl alcohol, potassium nitrate, and phosphoric acid were added each day as nutrients. Soda ash was added as required to control the wastewater pH. The initial wastewater parameters following start-up were: pH 8.1, temperature 8°C (47°F), dissolved oxygen (DO) 10.8 mg/1, nitrates (N) 14 mg/1, and phosphates (P) 4 mg/1. The biomass growth on the RBC media developed very slowly due to the low temperature. Sufficient alcohol, nutrients, and soda ash were added each shift to maintain the chemical oxygen demand (COD) between 100 to 200 mg/1, nitrates (N) 5 to 30 mg/1, phosphates (P) 1 to 5 mg/1, and the pH between 6.5 and 7.8. On the fourth day of operation the wastewater temperature increased to 12°C (53°F) and a very noticeable acceleration of biomass growth was observed.

The RBC operation continued in the batch mode until the eighth day, at which time an influent flow of 1.14 m^3/min (300 gpm) was started. The flow rate to the RBC units was steadily increased over the next few days up to 3 m^3/min (800 gpm).

pH FLUCTUATIONS

Shortly after start-up the facility encountered a period of pH fluctuations. During this period, the influent pH varied from 5.3 to 10.7 (figure 2). The variations in the influent pH were the result of a new pretreatment facility being unable to accurately control the acid feed rate for the pH control system. Until this problem was corrected, an attempt was made to adjust the pH in the equalization basin by the addition of soda ash; however, due to the absence of rapid mix equipment, this method was not completely successful.

As a result of these pH excursions, most of the biomass on the RBCs sloughed off. However, the biomass recovered without reseeding.

INITIAL OPERATION

During the next month of operation, despite the cold water temperature, additional pH excursions, and a highly

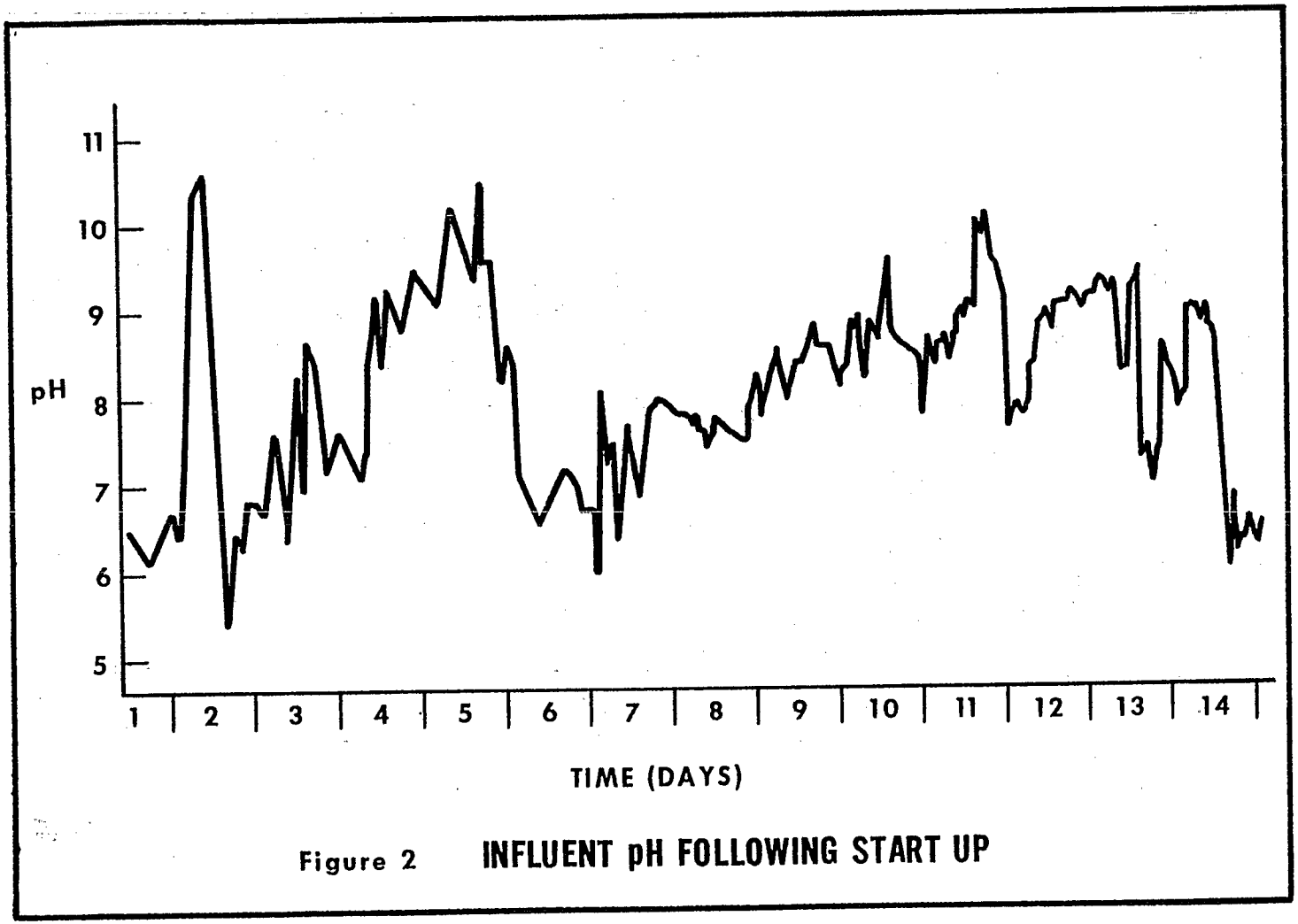

Figure 2 INFLUENT pH FOLLOWING START UP

variable organic load, the biomass growth continued to improve. The hydraulic loading to the RBC units was maintained constant except for adjustments necessary to maintain the equalization basin between the 60 to 90 percent level. However, wide variations in the organic concentration of the wastewater caused the COD removal efficiency to consistently fluctuate between 40 and 80 percent. Grab samples of the influent and effluent were collected each morning, five days a week. Figure 3 shows the COD of these samples collected during the second month of operation. The low influent CODs on Monday of each week were due to no manufacturing operations on weekends.

FULL FACILITY OPERATION

In an effort to provide better control of the organic loading to the RBCs, an on-line total organic carbon (TOC) analyzer was installed to monitor the RBC influent. The on-line TOC analyzer verified that, even with a one-day retention time in the equalization basin, sharp fluctuations in the organic concentrations of the RBC influent were occurring. A typical shock load caused an increase in TOC of 330 percent in less than three hours (figure 4). The results of the TOC analyses over a 23-day period are shown in figure 5. It was obvious from these data that a method of controlling the organic loading to the RBC was needed. The equalization basin was designed to be operated at between 60 and 90 percent capacity. Therefore, by maintaining the level of the basin at the low level when high organic loadings were expected and at the high level when low organic loadings were expected, the influent flow rate could be varied to minimize these fluctuations. A chart was prepared for use by the plant operators to control the organic loading to the RBC, based upon the influent flow rate and TOC value (figure 6). The instructions provided with the chart were as follows:

1. Keep the loading to the RBCs in the same loading zone whenever possible. Change the flow rate in small steps whenever it is necessary to change zones.
2. Decrease the basin level during periods of low organic loadings. Decrease the level to 60 percent on Sunday of each week.
3. Increase the basin level during periods of high organic loadings. Increase the level to 90 percent on Friday of each week.
4. The operators should maintain a record of the influent loadings by plotting the changes on the chart, using a new chart each day, and recording the time of changes.

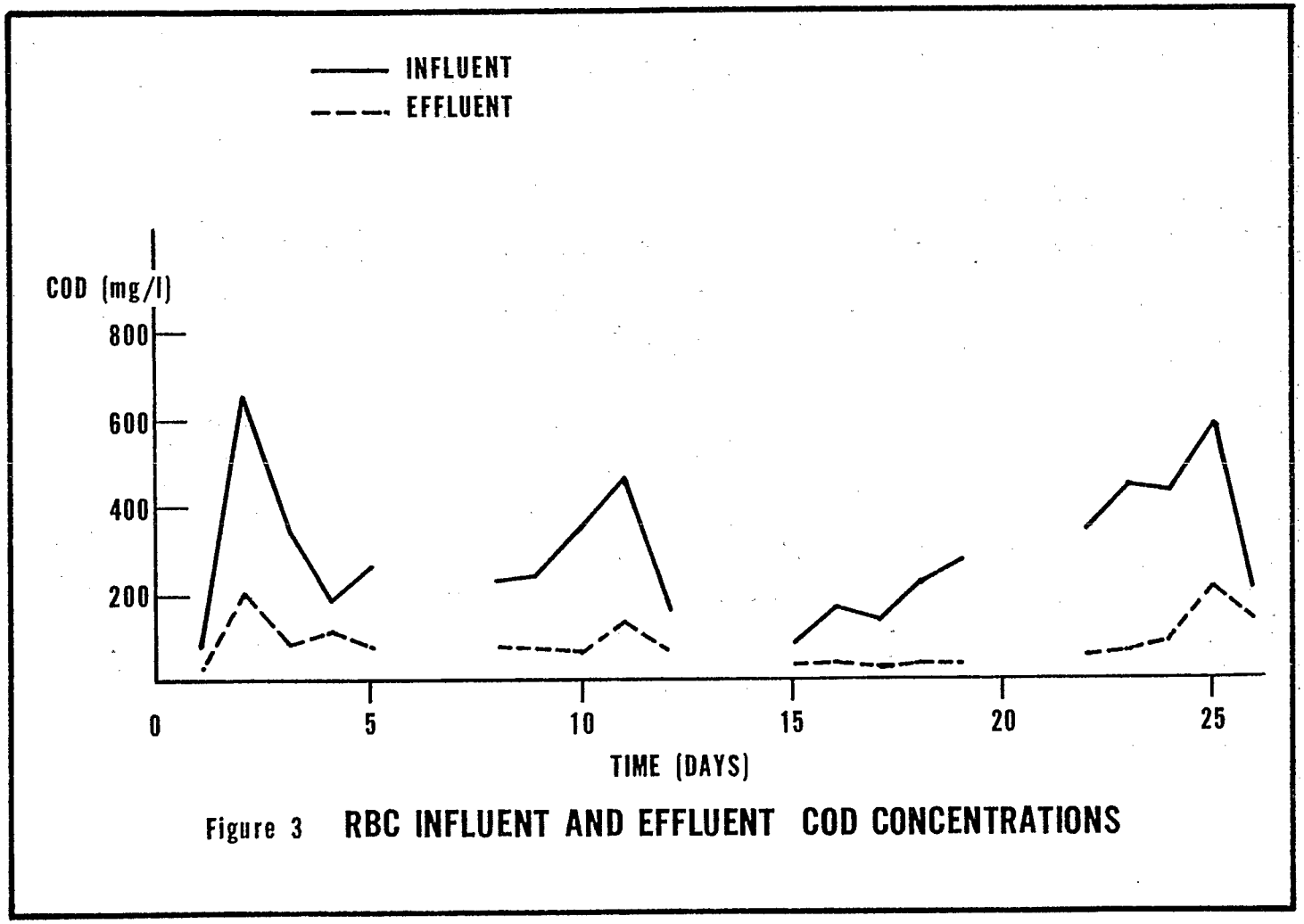

Figure 3 RBC INFLUENT AND EFFLUENT COD CONCENTRATIONS

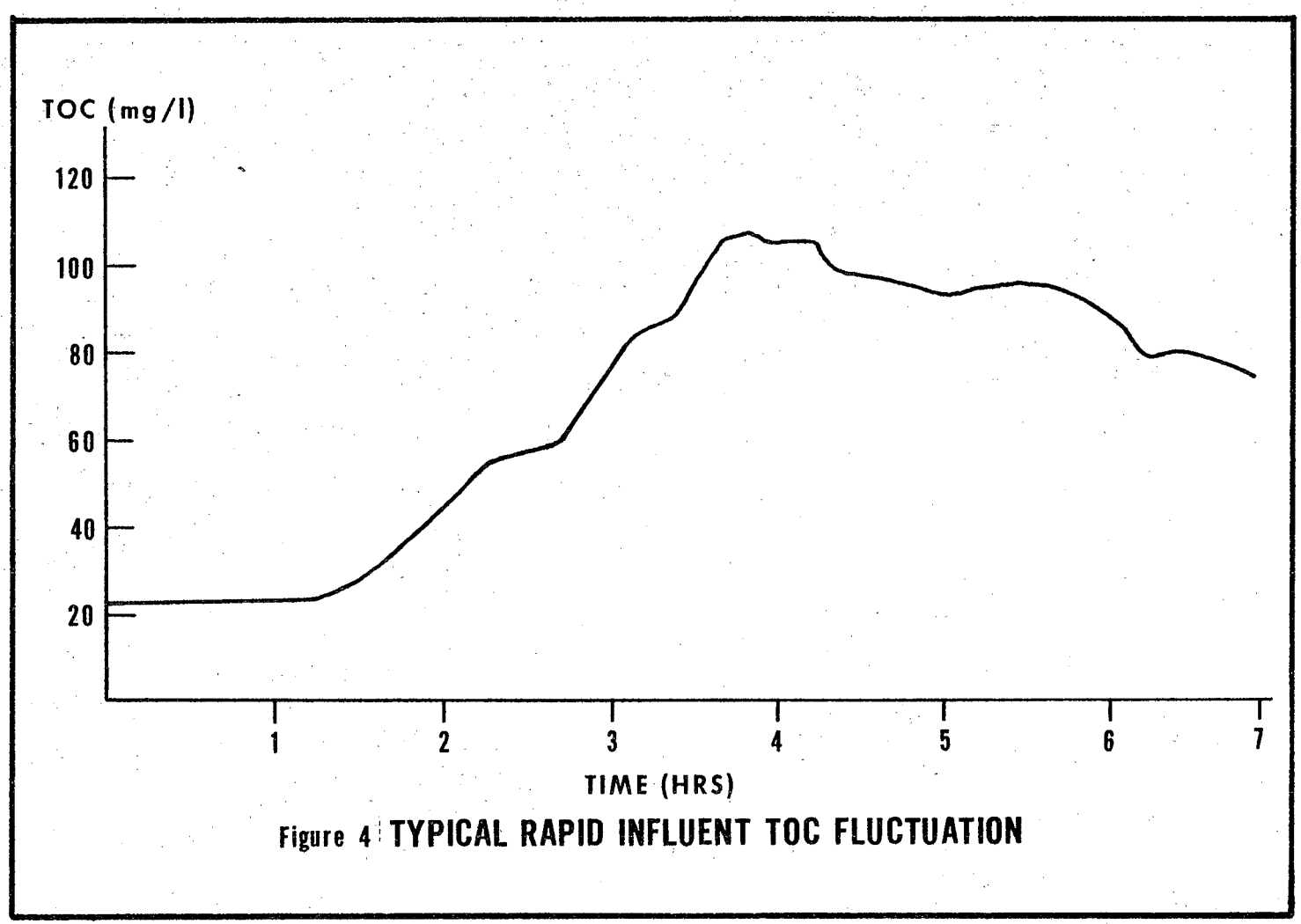

Figure 4 TYPICAL RAPID INFLUENT TOC FLUCTUATION

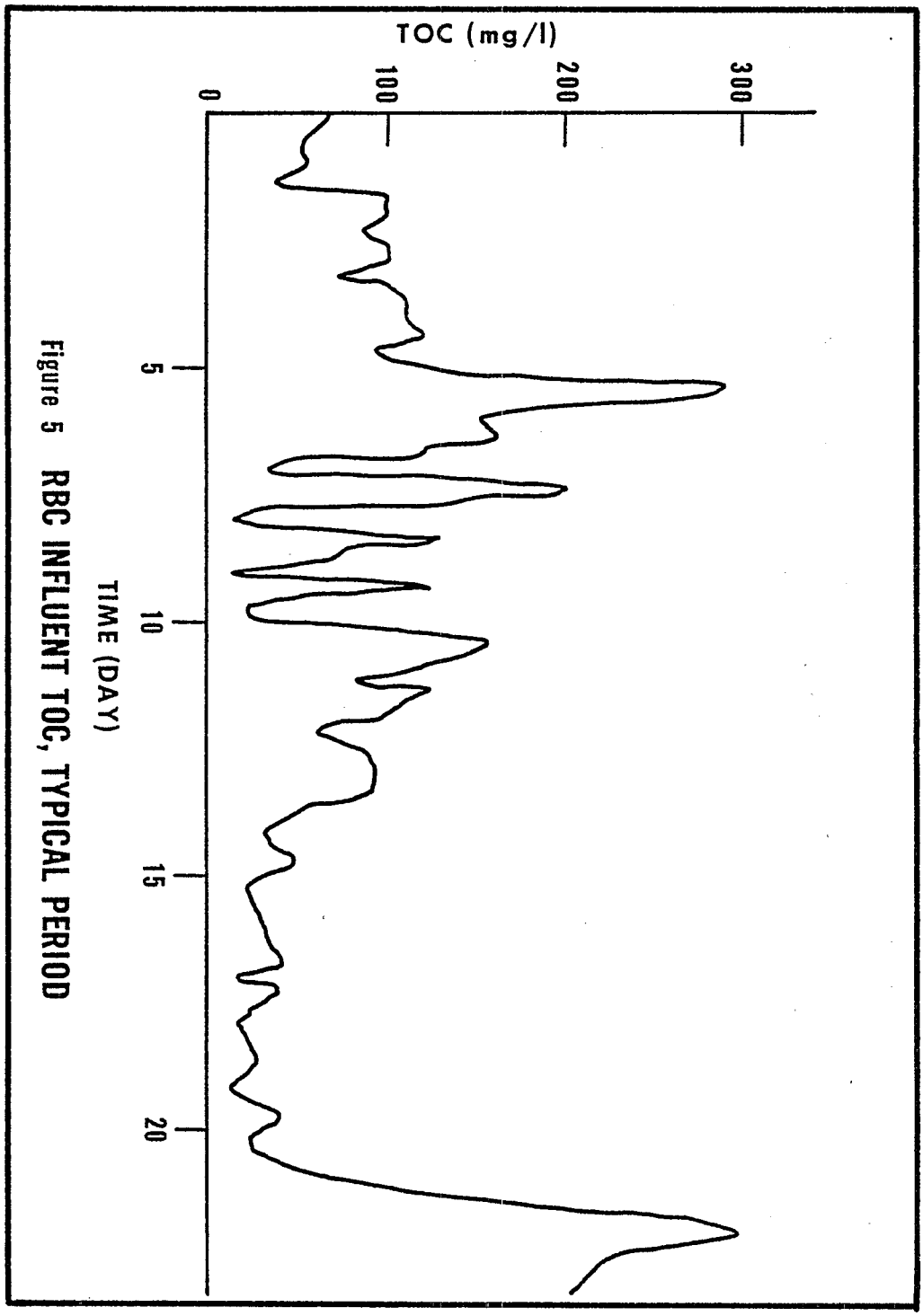

Figure 5 RBC INFLUENT TOC, TYPICAL PERIOD

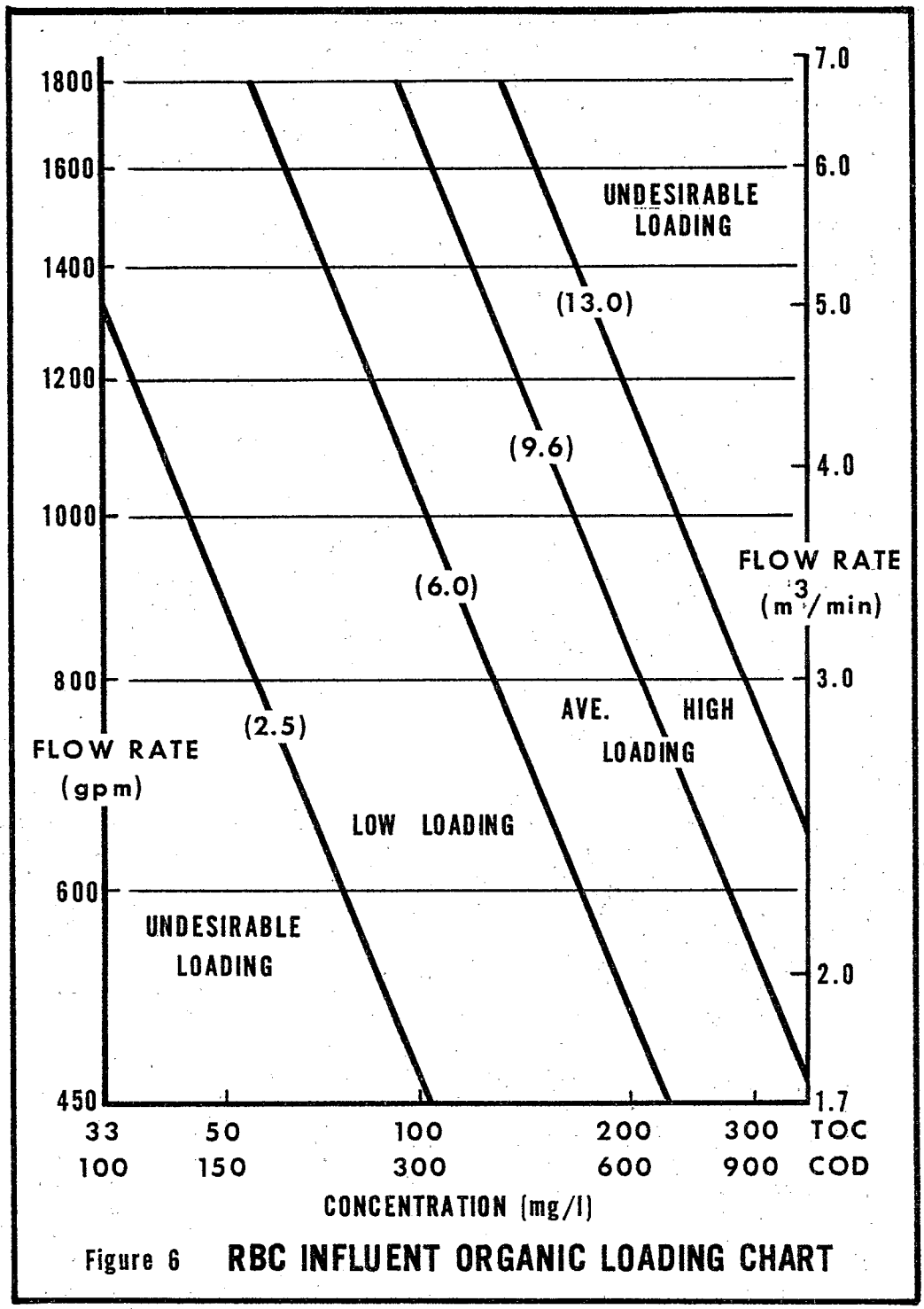

Figure 6 **RBC INFLUENT ORGANIC LOADING CHART**

The use of this chart and the TOC monitor helped control the organic loadings to the RBCs, but did not eliminate periodic shock loadings. Figure 7 shows the influent loading and effluent discharge for a two-month period. During this time, both of the RBC systems (eight shafts) were being used equally. The data for this chart were from analyses of grab samples collected at the start of day shift, five days a week. Twenty-four hour composite samples are collected once a week to verify compliance with the discharge permit. The results from these analyses show that the facility is meeting the discharge limitations.

The data from this same period were plotted as COD applied versus COD removed (figure 8) to determine the COD removal at various influent loadings. As can be seen from this graph, the data appear to be very scattered. A straight line represented by the equation $y = 0.83 x - 0.6$ was drawn on the graph to represent the normal operation of the RBC system. This corresponds closely to the equation developed during the pilot plant evaluation[1] of ($y = 0.83 x - 1.2$). This pilot plant equation was developed at steady state loadings which were varied over a period of time from 60 to 90 kg/1000 $m^2 \cdot d$ (12 to 18 lb/day/1000 ft^2). The data points on figure 8 falling considerably below this normal operation line are indications of stresses on the system due to shock loads.

PARTIAL FACILITY OPERATION

During the period of time the above data were collected, it can be seen from figures 7 and 8 that the RBC COD loadings were below the average design loadings of 37 kg/1000 $m^2 \cdot d$ (7.7 lb/day/1000 ft^2) most of the time. In order to evaluate the facility at design loadings, one RBC system (four shafts) was shut down in October 1981. The biomass growth on the RBC's media became heavier within the first few days of operation of only one system. The COD data from the grab samples, collected while only one system was operating, are shown in figures 9 and 10. The COD loadings during this period fluctuated as greatly as in the previous study; however, it can be noted from figure 10 that fewer data points fell considerably below the normal operating line. It appears that by increasing the organic loadings, thereby causing a heavier biomass, the system was more tolerant to shock loads. A detailed study has not been conducted to determine the effects of temperature on the organic removal efficiency. However, during the last 30 days of this study, the wastewater temperature varied from 5° to 10°C (42° to 50°F). It was noted by visual inspection

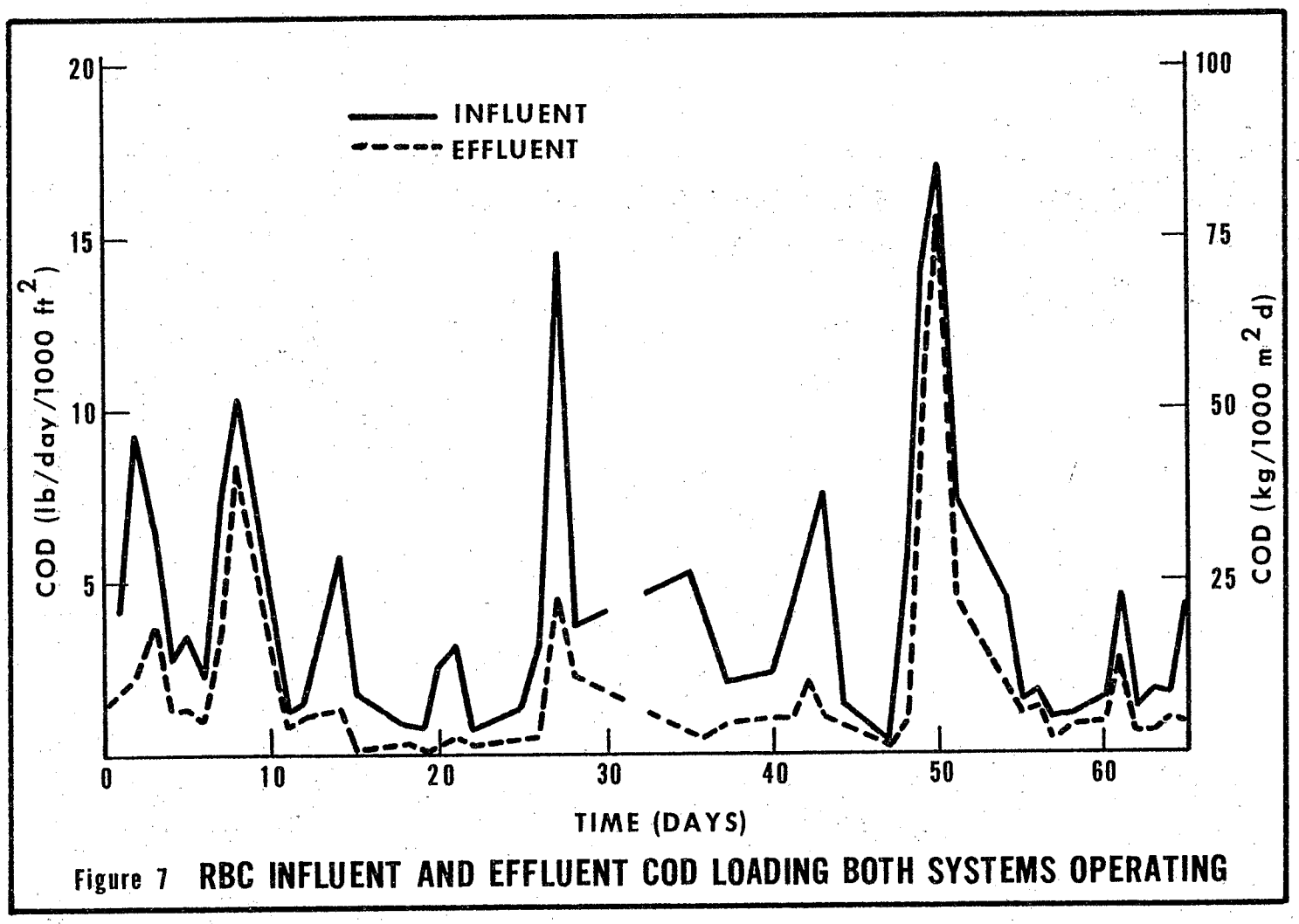

Figure 7 RBC INFLUENT AND EFFLUENT COD LOADING BOTH SYSTEMS OPERATING

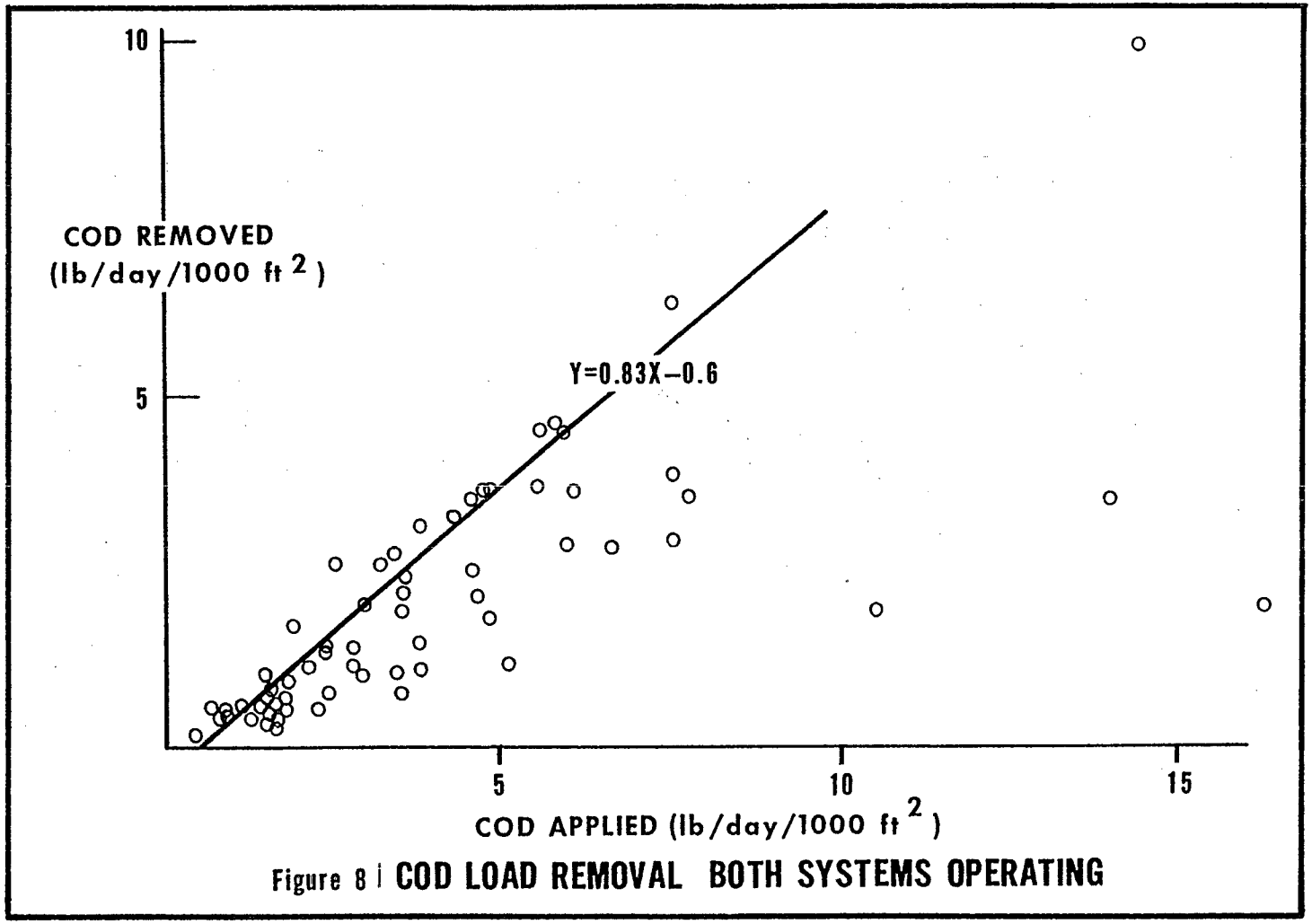

Figure 8 | COD LOAD REMOVAL BOTH SYSTEMS OPERATING

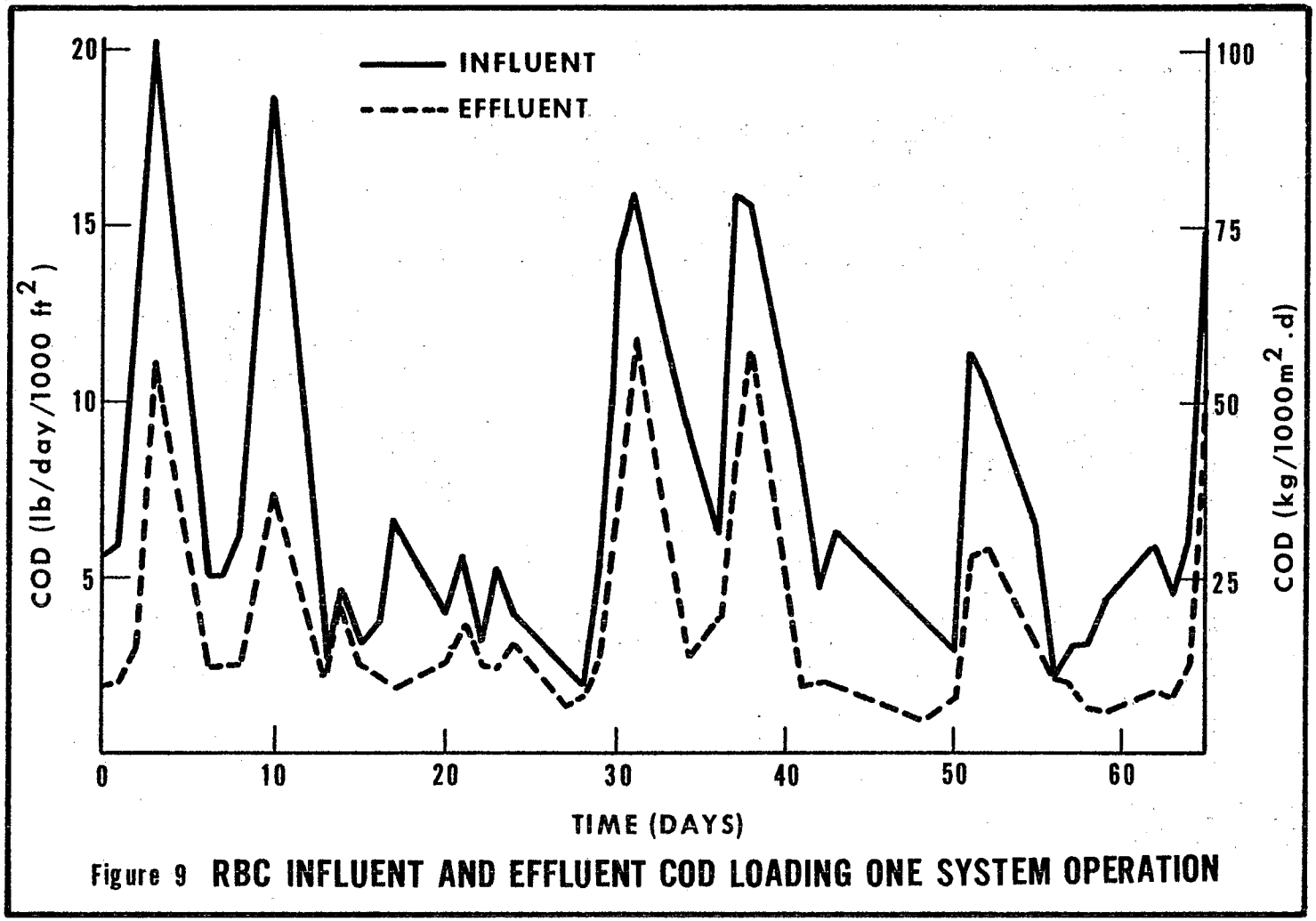

Figure 9 **RBC INFLUENT AND EFFLUENT COD LOADING ONE SYSTEM OPERATION**

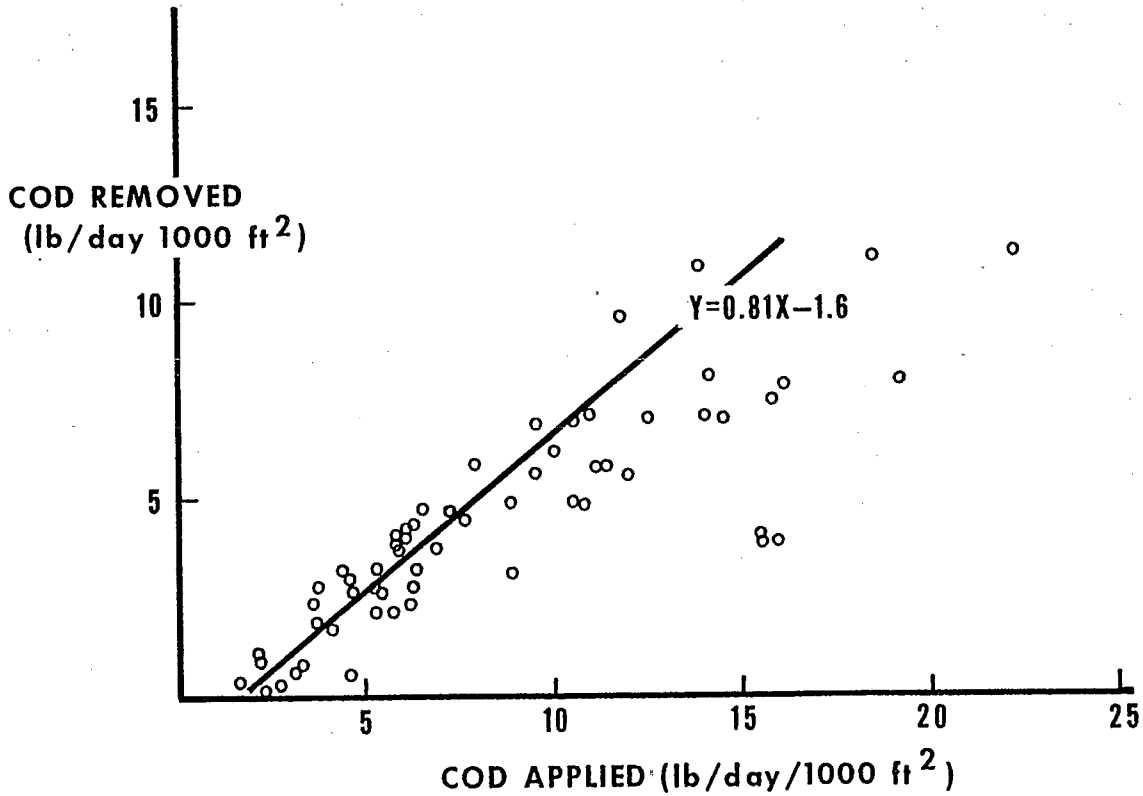

Figure 10 **COD LOAD REMOVAL ONE SYSTEM OPERATING**

that the colder wastewater caused the biomass growth to increass to compensate for the decreased activity of the biomass.

CONCLUSIONS

Based upon the one-year operation of the RBC facility on the wastewaters from this manufacturing facility, the following preliminary conclusions are drawn:

1. The biomass on RBC units can stand extreme stresses from pH, temperature, and shock organic loads without the system failing to a degree that it does not recover when the stress is removed.
2. A sudden increase in organic loading will not immediately increase the organic removal rate. However, the biomass growth will increase rapidly to increase the removal rate. The biomass growth rate will be directly related to the wastewater temperature because of the slower growth at lower temperatures.
3. Low wastewater temperature does not significantly decrease the organic removal efficiency at normal operating levels. However, at higher organic loadings, the RBC may not have the capacity for the additional biomass growth.

REFERENCES

1. Smith, L. L., and Zeigler, E., Jr., "Biological Treatment of a Munitions Manufacturing Facility Wastewater," Proceedings of the 33rd Industrial Waste Conference, Purdue University, May 1978, pp. 432-439.

TREATMENT OF STARCH INDUSTRIAL WASTE BY RBCs

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INTRODUCTION

Although successful treatment of industrial wastewaters by rotating biological contactors (RBC) has been earlier reported by many researchers, one important factor limiting the plant performance is the availability of oxygen in the system, in particular, when treating a high-strength organic waste (1-7). Oxygen transfer can become more effective by increasing the disc rotational speed but this application, however, is not practical because of: (a) high power consumption, and (b) high biomass slough-off induced by a high hydraulic shearing force on the surface of biodisc that reduces the overall treatment efficiency.

One alternative that may be able to treat high-strength organic wastes successfully under the normal operating conditions, is to replace the source of conventional air by pure oxygen. Bintaja et al. first employed oxygen in RBC system for treatment of Cheese waste (8). Later, Huang successfully treated synthetic milk waste by oxygen-enriched RBC system.

They concluded that use of pure oxygen in sufficient amount not only was able to improve COD removal satisfactorily but also increased sludge settleability(9).

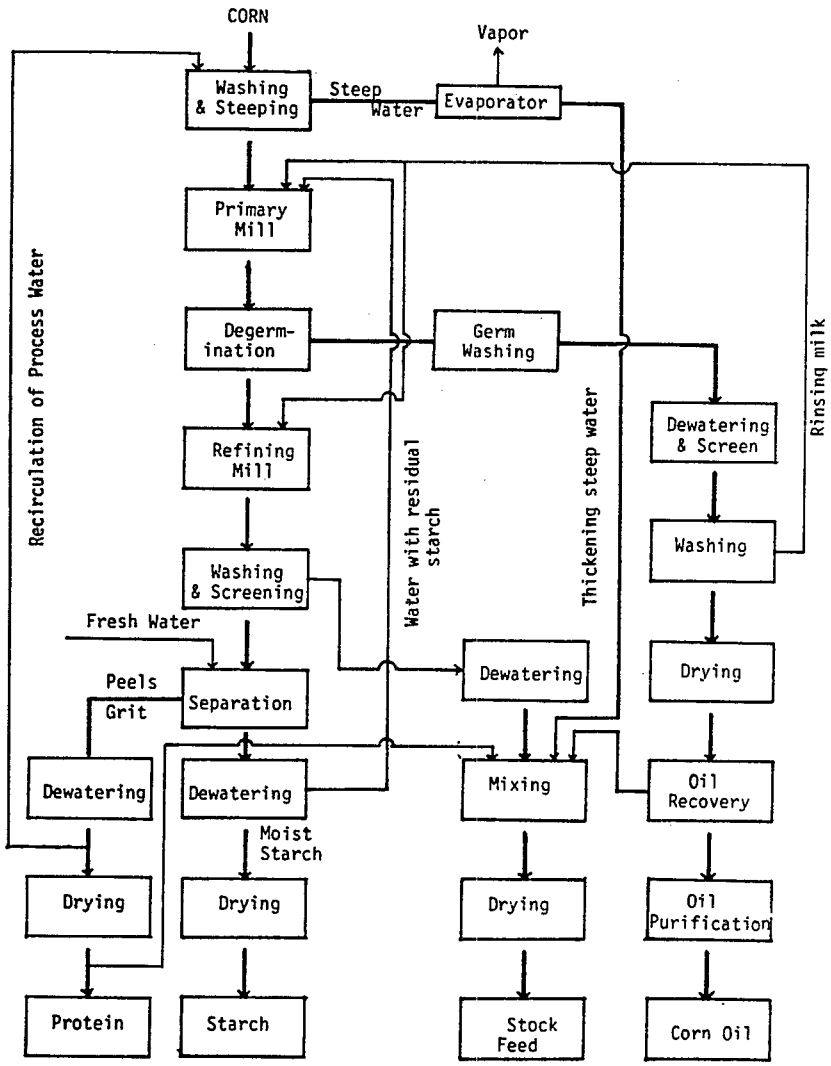
The present study was aimed to investigate the feasibility of using pure oxygen and conventional air RBC systems for the treatment of starch processing wastewater and also to study the effects of wastewater property, organic loading, pH, and dissolved oxygen concentration on RBC plant performance. The final goal of this research was to determine the kinetic data for future design of both RBC systems.

Starch Industry Wastewater. Starch consumption in the Republic of China (Taiwan) now reaches one hundred thousand tons annually, of which one half of this demand is supplied by local manufacturers and the other half is imported from the outside producers. As a result, Taiwan government has decided to assist private daily food industry in expanding the existing starch processing plants prior to 1984, so that the future domestic starch demand can be met.

The raw materials used presently for starch production in Taiwan are sweet potato and corn. Approximately 70% of the total starch is made of sweet potato because it is available locally. However, corn starch is more perferably used by consumers even the cost is higher. It is predicted that the use of corn for starch-manufacturing will increase considerably due to the current market demand.

An expansion of the existing starch processsing facilities is greatly concerned by environmental scientists and engineers in Taiwan. Because normally each ton of corn used for strach-making could produce 13.5 - 15.1 kg BOD and, more importantly, most of the starch manufacturing plants located in Tainan and Chiayi directly discharged their effluents into the small receiving streams. Purification of the plant effluent wastewaters is essentially necessary. The present study was initiated to investigate the treatment of corn strach manufacturer plant effluent by RBC systems. The flow diagram as presented later explains the process employed for corn starch manufacturing.

It can be seen in the flow diagram that although a closed-loop system is designed for the purpose of eliminating the discharge of process water, a considerable amount of wastewater is preduced due to leakage, overflowing, and accidential spill. Normally, a 0.4 - 3.0 m^3 wastewater generated per ton of corn used was found in a corn-starch manufacturing plant.



Flow Diagram of Corn Starch Plant

Biokinetics

The model employed for this study has been earlier proposed by Monod and Clark et al.(10,11). The basic equations are given as follows:

$$V(ds/dt) = Q\,S_o - Q\,S_e - (u_a/Y)\,A_w\,X_a - (u_s/Y_s)\,X_s V \qquad \text{----------(1)}$$

in which

V = reactor volume, m^3

ds/dt = rate of substrate removal, mg/l-sec

Q = waste flow, m^3/day

S_o = influent substrate concentration, mg/l

S_e = effluent substrate concentration, mg/l

u_a = specific growth rate of attached biomass, day^{-1}

u_s = specific growth rate of suspended biomass, day^{-1}

X_a = weight of attached biomass per unit disc surface area, g/m^2

X_s = concentration of suspended solids, mg/l

A_w = total disc surface, m^2

Y = yield coefficient of attached biomass (mass of biomass produced in kg/ mass of substrate removal in kg)

Y_s = yield coefficient of suspended biomass (mass of biomass produced in kg/ mass of substrate removal in kg)

In the fixed-film RBC system, substrate removal by attached biomass is much greater than suspended biomass. So, it is reasonable to assume that the term $(u_s/Y_s)\,X_s\,V$ in Eq. 1 can be eliminated. And then Eq. 1 becomes

$$v(ds/dt) = Q\,S_o - Q\,S_e - (u_a/Y)\,A_w\,X_a \qquad \text{-------(2)}$$

According to Monod, the change in biomass under substrate limiting condition can be expressed as

$$\mu = u_m \{ S/ (K_s + S)\} \quad \text{---------} \quad (3)$$

in which

u_m = maximum specific growth rate, day^{-1}
K_s = half saturation constant, mg/l
S = limiting substrate concentration, mg/l

Eqs 2 and 3 can be combined under a steady-state condition (ds/dt)= 0 and expressed in linear form as shown below.

$$1/\lambda = (K_s/\beta)\ (1/S^*) + 1/\beta \quad \text{---------} (4)$$

in which

$\lambda = (Q/A_w)(S_o - S^*)$, mg/m^2-day
$\beta = (u\ /Y)\ X_a^*$, mg/l-day
S^* = effluent substrate concentration at steady-state condition, mg/l
X_a^* = attached biomass concentration at steady-state condition, mg/l

The values of K_s and β can be calculated by plotting $1/\lambda$ versus $1/S^*$. The ordinate intercept and slope of the line are equal to $1/\beta$ and K_s/β, respectively.

Materials and Methods

The chemical composition of the starch processing wastewater employed for this study is shown in Table 1. Apparently, organic carbon concentration as BOD or COD and solids content were high and the wastewater pH was in the acidic condition.

Throughout the entire study, the BOD, COD, pH, DO were monitored in each stage of the RBC system. The laboratory procedures used to determine the above mentioned parameters are specified in "Standard Methods" (12). The % oxygen content in feed gas was measured in each stage of the pure oxygen RBC

system by Beckman Oxygen Analyzer.

Table 1. General Property of the Wastewater

Parameter	Concentration, mg/l
BOD	2,700-3,900
COD	5,200-7,100
Organic-N	300-400
Total Solids	3,300-5,400
Total Volatile Solids	2,900-4,700
pH	4.05-4.55
BOD/N ratio	10: 1
N/P ratio	5: 1

Two RBC pilot plants were constructed identically in size. The main difference between them was that the biodiscs in one system were exposed to air and in other system they were constantly contacted with pure oxygen. The physical structure of these pilot plants is described in Table 2.

A total of 47% disc area was submerged and all discs were supported by a common shaft in each stage. Three shafts in each system were driven by the same motor so that they were rotated at the same velocity. However, this motor enabled to rotate at the speed varying from 0 to 200 rpm.

To produce an airtight condition for pure oxygen RBC system, a plexiglass cover was bolted down at the top of each biological reactor. The shaft was made of a ½" steel rod that was connected to motor drive system through the shaft holes at where ½" O-rings were poisitioned. Replacement of O-rings was necessary when they were worm down from friction. The re-

Table 2. Pilot Plant Specifications

	Stage Number		
(1) Biodiscs	1st	2nd	3rd
Diameter, cm	26	26	26
Thickness, cm	0.2	0.2	0.2
Spacing, cm	4	2.8	2.2
Number	18	25	31
Effective area, m^2	1,894	2,630	3,261
% Submerged area	47	47	47
(2) Reactor			
Material	Plexiglass Plate		
Cross Sectional Area, cm^2	260 x 405		
Length, cm	760	760	760
Volume, liters			
Gross	25.27	25.27	25.27
Net	24.40	24.00	23.70
(3) Enclosure*			
Material	plexiglass cover		
Size, cm	760 x 260 x 150		
Volume, liters	39	38.6	38.2

* in the pure oxygen system only.

actor structure is presented in Figure 1 in detail. All experiments were performed at a room temperature = 28 ± 2°C.

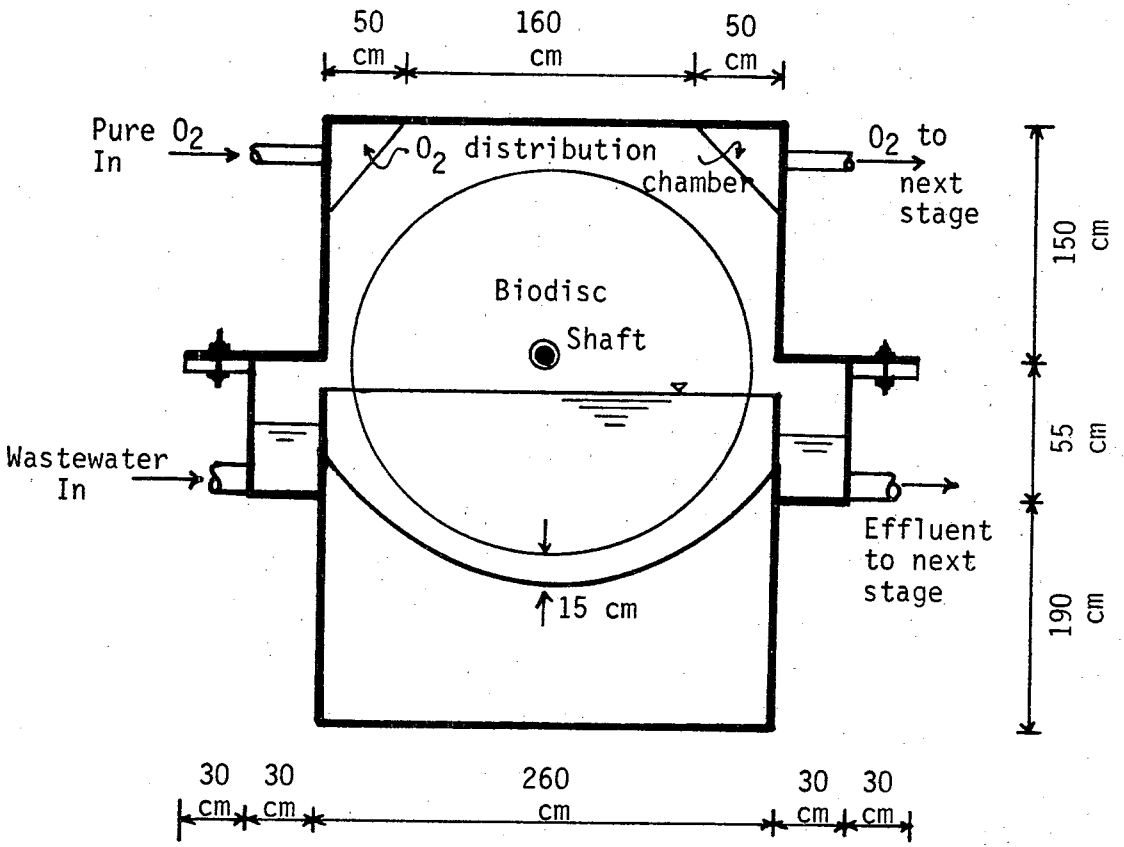


Figure 1. Structure of Pure Oxygen RBC

Results and Discussion

To study the feasibility of using the RBC systems for the treatment of starch manufacturing wastewater, the effects of influent substrate concentration, organic loading, DO, pH, disc rotational speed (RS), and hydraulic retention time (RT)on substrate removal efficiency were investigated.

(1). Disc Rotational Speed. Antonie earlier suggested that the rotational speed of biodiscs should be controlled to exceed one foot per second (18.3 m/min.) so that oxygen limitation could be avoided (8). The effect of disc rotational speed on treatment efficiency was tested under the influent substrate concentration ranged from 370 mg/l to 1,300 mg/l as BOD and from 860 mg/l to 2,900 mg/l as COD. The rotational speed was controlled at 15, 20, and 25 rpm, respectively.

Table 3 summarized the results of this study conducted in the pure oxygen RBC system. It was found that for all substrate conditions presently investigated, neither BOD nor COD removal was significantly increased due to the increase in disc rotational speed. In other words, there is no need to maintain the rotational speed over 15 rpm (12.3 m/min.), according to the present study.

Table 3. Effects of Influent Substrate Concentration and Disc Rotational Speed on % COD and BOD Removal

Influent COD Conc. (mg/l)	% Reduction		
	Rotational Speed, rpm		
	15	20	25
860	91.9	91.4	91.6
1,580	94.6	94.3	94.0
1,900	93.0	93.6	93.0
2,300	88.3	90.7	91.0
2,900	87.2	87.8	88.3

Table 3. Continued

Influent BOD Conc. (mg/l)	% Reduction		
	Rotational Speed, rpm		
	15	20	25
370	99.1	99.2	99.0
650	98.9	98.7	98.5
900	98.0	98.2	98.4
1,000	97.1	98.0	98.0
1,300	96.2	96.5	96.8

(2). pH Level. To obtain an optimum treatment efficiency, the acidic starch wastewater was first neutralized by NaOH and then mixed with KH_2PO_4 solution to increase buffering capacity. The pH level after adjustment was controlled to be within the range of 8.1 to 8.9. Because of the chemical nature of the wastewater, the pH was reduced approximately one unit during 12 hours storage. Normally, the higher the organic content in the wastewater the faster the pH drop. For this reason, the pH of the feed solution was adjsuted in accordance with COD concentration. Table 4 shows the results of pH changes in the feed as well as in each stage of the RBC systems.

The data as seen in Table 4 indicated that in most cases the reduction of pH occurred in the first stage of the system and an increase was found in the subsequent stages. The pH drop was probably due to the accumulation of organic acids resulted from active decomposition of organic substances whereas the pH increase was induced by CO_2 production, a metabolic by-product in this case. Additionally, the present study shows that the influence of substarte concentration, rotational speed, and retention time on pH was not significant.

It is evident from this study that the buffer solution added to the wastewater is adequate to maintain a proper pH between 7.0 and 7.6. However, it is important to investigate whether the chemical addition of phosphate buffer can be reduced, and to determine its minimum requirement. Further study is essentially needed for the future operation of the RBC system.

Table 4. Effect of Influent COD, Disc Rotational Speed, and Hydraulic Retention Time on pH

RBC System	Operating Condition			pH Value Stage Number			
	COD mg/l	Rotational Speed, rpm	Retention Time, hrs	Feed*	1st	2nd	3rd
Oxygen	1,000	15	12	8.3/6.5	7.1	7.2	7.4
	2,000	15	12	8.5/6.7	7.3	7.4	7.5
	3,000	15	12	8.8/6.7	7.2	7.3	7.5
	4,000	15	12	8.8/8.3	7.0	7.1	7.2
	5,000	15	12	8.9/7.8	7.1	7.5	7.6
	1,000	15	6	8.2/7.1	6.9	6.9	7.0
	2,000	15	6	8.3/7.6	6.9	7.1	7.3
	3,000	15	6	8.5/7.6	6.9	7.1	7.2
	4,000	15	6	8.7/7.3	6.7	7.1	7.4
	5,000	15	6	8.9/7.9	7.0	7.1	7.4
	1,000	15	4	8.1/7.4	7.6	7.6	7.6
	2,000	15	4	8.4/7.3	7.1	7.2	7.3
	3,000	15	4	8.6/7.5	7.1	7.3	7.6
	4,000	15	4	8.8/8.0	7.0	7.1	7.3

Table 4. Continued

RBC System	Operating Condition			pH Value			
	COD mg/l	Rotational Speed, rpm	Retention Time, hrs	Stage Number			
				Feed*	1st	2nd	3rd
Oxygen	2,000	20	12	8.4/6.4	7.4	7.3	7.5
	3,000	20	12	8.9/6.8	7.3	7.4	7.6
	4,000	25	12	8.9/7.2	7.5	7.6	7.7
	5,000	25	12	8.9/8.4	7.1	7.2	7.2

* (initial pH/pH after 12 hrs storage)

(3). Oxygen Consumption and DO Level. The effect of organic loading on oxygen consumption and DO level under different disc rotational speeds (15, 20, and 25 rpm) was studied for both air and pure oxygen RBC systems. The results are shown in Table 5.

A. In Pure Oxygen RBC System. The feed gas contained 99.5% oxygen and the feed rate was constantly controlled at 1,000 cc/min.. Table 5 clearly shows that the % O_2 gas remaining and the % O_2 utilization decreases as the number of RBC stages increases. But the influence of organic loading on oxygen utilization is different, that is the % O_2 consumption increases or the % O_2 remaining decreases when the organic loading increases. These results are expected because: (a) the removal of organic substrate takes place rapidly in the first stage of the RBC system that consumes more oxygen, and (b) in addition to the oxygen uptake in each stage, the accumulation of CO_2 reduces the % O_2 remaining in the feed gas.

The DO level in the oxygen RBC system was highly affected by the influent substrate condition. In two cases presently investigated, the DO level reached zero in the first stage as the influent COD and organic loading exceeded 5,000 mg/l and 95 g. COD/m^2-day, respectively, in accordance with Table 6. Additionally, it was also found that the DO level increased as the number of RBC stages increased.

Table 5. Effect of Substrate Concentration and Loading On Oxygen Utilization In Pure Oxygen RBC System *

Rotational Speed rpm	Influent COD mg/l	COD Loading g/m^2-day	$\%O_2$ Remaining Stage Number			$\%O_2$ Utilization Stage Number		
			1s	2nd	3rd	1st	2nd	3rd
15	1,000	19	92.4	91.0	90.3	7.1	1.4	0.7
20	1,000	19	92.8	91.8	90.8	6.7	1.7	1.0
15	2,000	38	88.6	86.8	86.2	10.9	2.0	0.6
20	2,000	38	89.8	88.1	87.4	9.7	1.7	0.7
25	2,000	38	89.6	87.1	86.6	2.5	2.5	0.5
15	3,000	57	86.0	84.1	83.2	13.5	1.9	0.9
20	3,000	57	86.8	85.0	83.9	12.7	1.3	1.1
25	3,000	57	86.6	84.3	83.6	12.9	2.3	0.7
15	4,000	76	85.4	83.7	82.5	14.1	1.7	1.2
15	5,000	95	84.0	81.4	79.0	15.5	2.6	1.6

* A 99.5% oxygen feed gas was injected into the first stage of the RBC system

Table 6. Effects of Substrate Concentration and Loading on DO Level in Oxygen and Air RBC Systems

Influent COD mg/l	COD Loading g/m^2-day	Oxygen RBC[1] stage number			Air RBC[2] stage number		
		1st	2nd	3rd	1st	2nd	3rd
1,000	19	8.5	19.3	21.6	4.6	8.0	8.6
2,000	38	6.0	20.3	18.0	1.6	4.0	7.5
3,000	57	4.2	13.0	16.0	0	1.5	5.0
4,000	76	1.5	9.8	12.0	0	0	0
5,000	95	0	6.0	9.0	0	0	0
7,000	133	0	0.2	0.5	0	0	0

1. Disc rotational speed= 15 rpm
2. Disc rotational speed= 20 rpm

B. In Air RBC System. The effect of substrate concentration and loading on DO level in the conventional air system is similar to the pure oxygen system. However, the DO levels for all three stages in the air RBC system were considerably lower. An anaerobic condition was found in the first stage of the system when the organic loading was equal to 57 g. COD/m^2-day and in all stages when the organic loading exceeded 76 g. COD/m^2-day.

Biological response to oxygen deficiency in RBC systems shows no significant difference from the other types of wastewater treatment processes. Two interesting evidences have been observed, one is the conversion of sulfate compounds

Table 7. Effect of RBC Stage On Substrate Removal Under Various Organic Loadings and Retention Times

1. Air RBC System

Influent Conc. mg/l		Organic* Loading	Retention Time	% Removal**			
				Number of Stage			
BOD	COD	g/m^2-day	hrs	1st	2d	3rd	overall
340	760	6.5/14.4	6	88.8/82.4	7.0/ 3.1	1.6/ 3.0	97.4/89.0
450	820	8.6/15.4	6	91.6/82.9	7.0/ 2.9	0.3/ 1.6	98.9/87.4
480	865	9.1/16.4	6	90.7/82.4	6.2/ 3.1	1.9/ 1.7	98.8/87.3
1,090	1,980	20.7/36.1	6	86.7/63.1	5.0/16.6	2.9/ 5.8	94.0/85.6
1,120	2,010	21.3/38.2	6	84.8/61.2	6.4/17.4	2.1/ 6.7	93.3/85.3
1,335	2,380	25.4/45.2	6	57.8/52.5	27.9/26.1	6.1 6.4	91.0/85.1
1,470	2,640	27.9/50.2	6	56.5/50.0	28.2/24.0	4.4/10.7	89.1/84.8
1,800	3,280	34.2/62.3	6	44.4/29.9	22.7/31.0	16.0/17.6	83.1/78.6
2,350	4,300	44.7/81.7	6	33.6/21.4	17.8/21.8	15.4/18.8	66.8/62.1
2,900	5,195	55.1/98.7	6	30.7/18.8	19.3/21.8	5.1/ 11.2	55.2/51.8

*(BOD Loading/COD Loading)

** (% BOD Removal/% COD Removal)

Table 7. Continued

2. Pure Oxygen RBC System

Influent Conc. mg/l BOD	COD	Organic* Loading g/m^2-day	Retention Time hrs	% Removal** Number of Stage 1st	2nd	3rd	Overall
265	480	15.1/ 27.4	4	45.3/39.6	27.6/31.0	28.6/62.5	92.5/84.4
650	1,190	37.1/ 67.8	4	47.7/34.5	23.5/28.2	61.5/37.5	84.6/70.6
1,160	2,080	66.1/118.6	4	32.8/27.9	28.2/22.7	51.8/31.0	76.7/61.5
1,270	2,370	72.4/135.1	4	29.1/22.8	23.3/25.7	52.9/35.3	74.4/62.9
1,495	2,780	85.2/158.5	4	29.8/26.3	25.7/22.4	29.5/24.5	63.2/56.8
1,930	3,560	110.0/202.9	4	18.7/14.1	15.3/15.7	30.0/24.0	51.8/44.9
2,200	4,130	125.4/235.4	4	15.9/11.6	13.5/ 9.6	26.9/24.2	46.8/39.5
2,500	4,540	142.5/258.8	4	12.0/ 9.7	15.5/12.2	24.7/22.2	44.0/38.3
550	980	20.9/ 37.2	6	55.5/51.0	71.4/68.8	50.0/23.3	93.6/88.3
820	1,490	31.2/ 56.6	6	56.1/52.7	72.2/68.8	50.0/34.1	93.0/90.3
1,020	1,790	38.8/ 68.0	6	22.1/25.1	78.2/72.0	60.0/46.7	93.3/88.8
1,070	1,890	40.7/ 71.8	6	21.5/20.6	50.0/48.7	82.4/68.9	93.1/87.3
1,690	3,100	64.2/117.8	6	39.6/34.2	33.3/22.1	57.4/40.3	89.8/69.4
1,900	3,500	72.2/133.0	6	31.6/25.1	45.8/31.3	36.2/31.7	76.3/64.9

Table 7. Continued

2. Pure Oxygen RBC System

Influent Conc. mg/l BOD	COD	Organic Loading* g/m^2-day	Retention Time hrs	% Removal** Number of Stage 1st	2nd	3rd	Overall
2,100	3,870	79.2/147.0	6	28.6/24.6	37.3/29.5	41.5/32.0	73.8/63.8
2,790	4,900	106.0/186.2	6	24.7/16.3	28.6/25.6	28.0/26.8	61.3/54.5
2,820	5,030	107.2/191.9	6	19.2/15.5	32.0/23.5	27.1/20.9	59.9/48.9
3,990	7,100	151.6/269.8	6	12.3/11.7	28.6/45.0	28.0/55.4	54.9/47.2
370	864	7.0/ 16.4	12	98.4/87.9	50.0/27.6	0.0/ 7.9	99.2/91.9
400	900	7.6/ 17.1	12	96.3/85.9	66.7/35.4	40.0/17.1	99.3/92.4
675	950	12.8/ 18.1	12	97.0/86.3	40.0/29.2	41.7/11.8	99.0/92.1
825	1,580	15.7/ 30.0	12	97.9/91.5	29.4/28.9	25.0/12.5	98.9/94.6
840	1,925	16.0/ 36.6	12	96.3/91.2	38.9/14.7	26.3/ 6.9	98.3/93.0
960	1,940	18.2/ 36.9	12	95.3/88.1	28.9/34.8	18.8/ 7.3	97.3/92.8
1,020	2,310	19.4/ 43.9	12	94.9/81.4	30.8/25.6	19,4/15.6	97.2/88.3
1,920	3,500	36.5/ 66.5	12	59.4/80.2	62.8/28.1	34.5/17.7	90.1/86.1
1,980	3,660	37.6/ 69.5	12	57.1/48.6	62.4/59.7	37.5/34.0	89.9/85.6
2,960	5,400	56.2/102.6	12	30.4/48.1	59.5/57.9	46.8/35.6	88.9/79.6
3,900	7,100	74.1/134.9	12	23.1/15.4	55.0/55.4	48.1/46.1	85.0/78.3

to an offensive H_2S gas and the other is the overgrowth of filamentous microorganisms that reduces the sludge settleability. Unfortunately, no identification of thread-like organisms was obtained from the present study.

(4) COD and BOD Removal As a Function of Substrate and Loading Conditions. The effect of influent substrate concentration and organic loading on RBC plant performance was investigated at the disc rotational speed equal to 15 and 20 rpm, respectively, for the conventional air and pure oxygen system. The results of the study were summarized in Table 7.

It is apparent from Table 7 that the % COD or BOD removal is highly dependent upon the system operating conditions. In general, the lower the influent COD or BOD concentration and organic loading the higher the % substrate removal. The permissible organic loading and retention time for 90% BOD and 80% COD removal are estimated for both RBC systems as follows:

System	% Removal		Retention Time, hrs	Organic Loading $g./m^2$-day	
	BOD	COD		BOD	COD
Oxygen	90	80	4	19.0	49.0
	90	80	6	31.0	70.0
	90	80	12	38.0	110.0
Air	90	80	6	28.0	62.3

To meet the above requirements, the oxygen RBC system was perferably operated under a retention time between 6 and 12 hours. At the retention time of 6 hours, the oxygen RBC system can be loaded at approximately 9.6% on COD basis and 11% on BOD basis higher than the conventional air system. Additionally, it was also found that in the oxygen system, the organic loading rate at 6 hours retention time was 30% and 38% higher than at 4 hours, respectively, for 80% COD and 90% BOD removals.

The net % reduction of COD and BOD from each stage of the RBC systems were also reported in table 7. It was observed that when the organic laoding was low (i.e., BOD load < 19.6 $g./m^2$-day and COD load < 43.9 $g./m^2$-day), effective

removal of BOD and COD was obtianed in the first stage of RBC. The subsequent stages become important as a result of increasing the organic loading or shortening the hydraulic retention time. The overall treatment efficiency as a function of BOD or COD loading is illustrated in Figures 2 and 3.

The relationship between organic loading and substrate removal per unit surface area per time is shown in Figures 4 and 5. A linear relationship was found for the conventional air system when the organic loading was below 60 g. BOD or COD/ m^2-day, and for the pure oxygen system as the loading was not in excess of 70 g. BOD or COD/m^2-day. The equations used to describe the above mentioned relationship are given as

Air RBC System:

$$\Phi = 0.8 + 0.82\ \omega, \text{ if } \omega < 60\ g/m^2\text{-day} \quad \text{--------(5)}$$

Oxygen RBC System:

$$\Phi = 2.8 + 0.79\ \omega \quad \text{if } \omega < 70\ g/m^2\text{-day} \quad \text{--------(6)}$$

in which

Φ = Organic Removal in g COD or BOD removed/ m^2-day

ω = Organic Loading in g COD or BOD/m^2-day

Figures 4 and 5 also show that the predicted organic removal efficiency Φ calculated from Eqs. 5 and 6 exceeds the observed values when the organic laoding ω was greater than the limits already mentioned.

(5) Foaming and Brdiging. Throughout the entire study, foaming never occurred in the pure oxygen RBC system but it was a serious problem in the conventional air RBC system (see Figures 6 and 7). It was found that the first stage of the conventional air system was covered by foam at the organic loading equal to 38 g COD/m^2-day. When the loading was increased to exceed 50.2 g. COD/m^2-day, the operation of both first and second stages of the RBC system was interferenced. The low efficiency of the conventional air RBC system could be partially caused due to the foaming that resulted in a low oxygen transfer to attached biomass.

The maximum thickness of the biofilm was approximately equal to 2.5 cm on the first-stage biodiscs of the pure oxygen

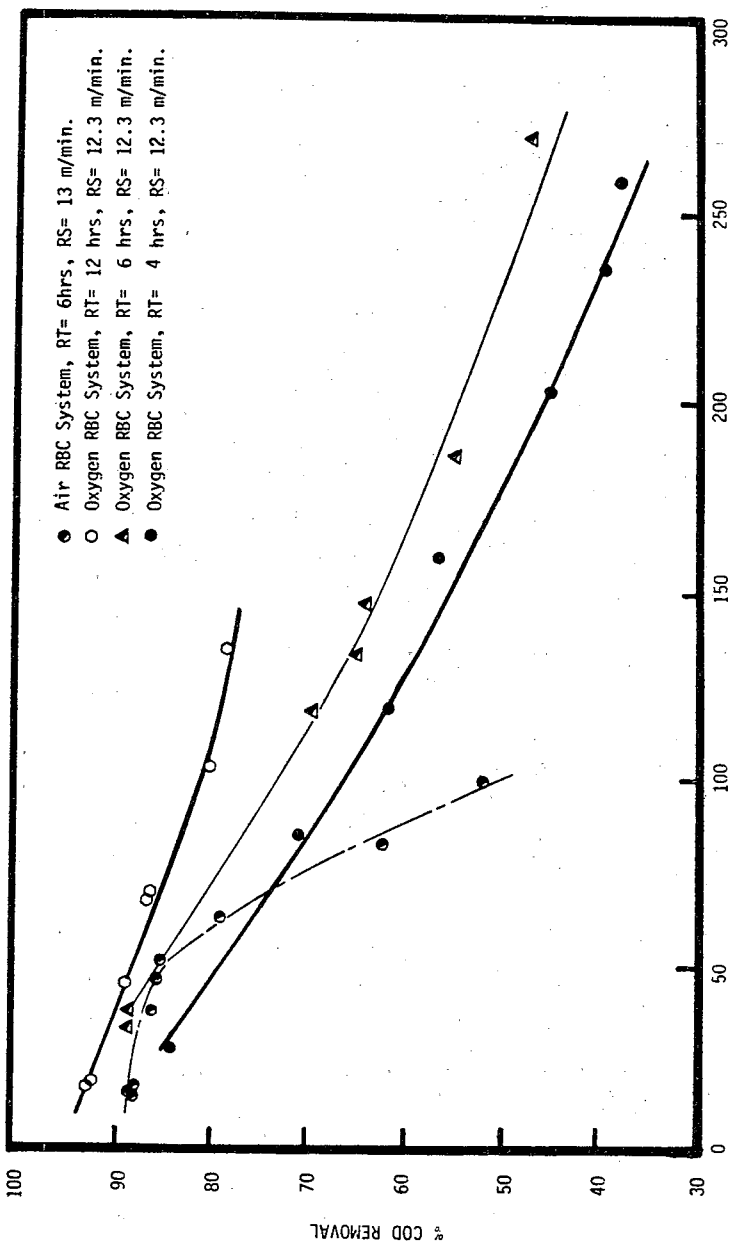


Figure 2. Effect of COD Loading on % COD Removal

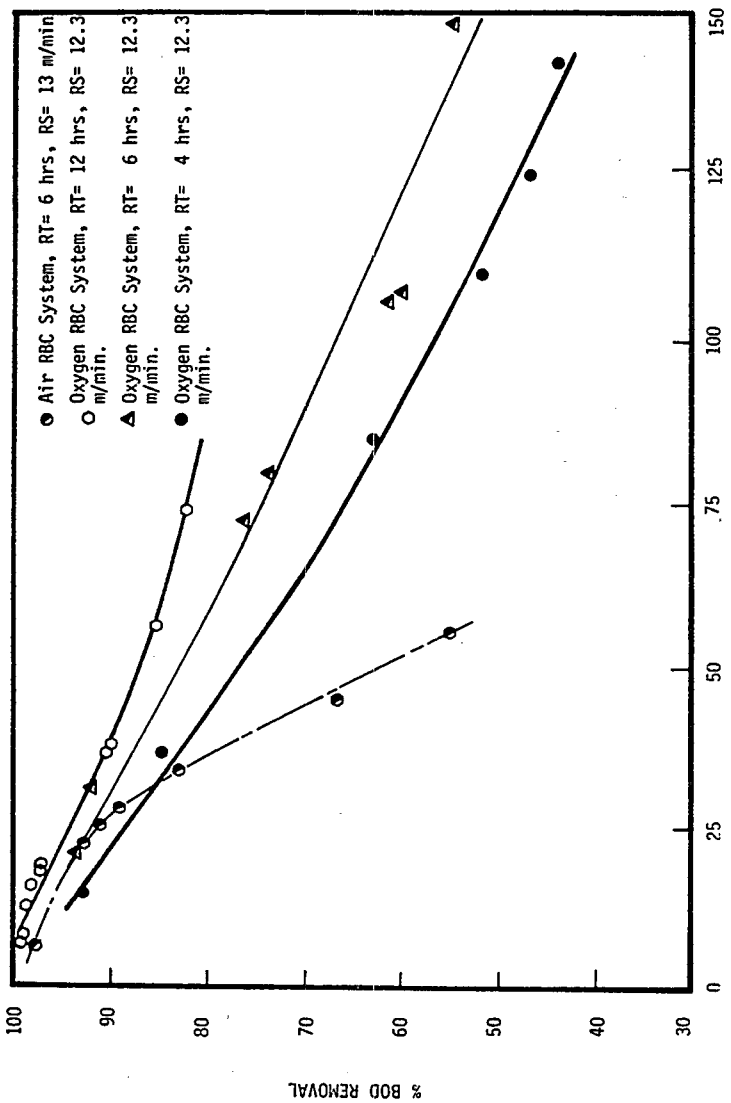


Figure 3. Effect of BOD on % DOD Removal

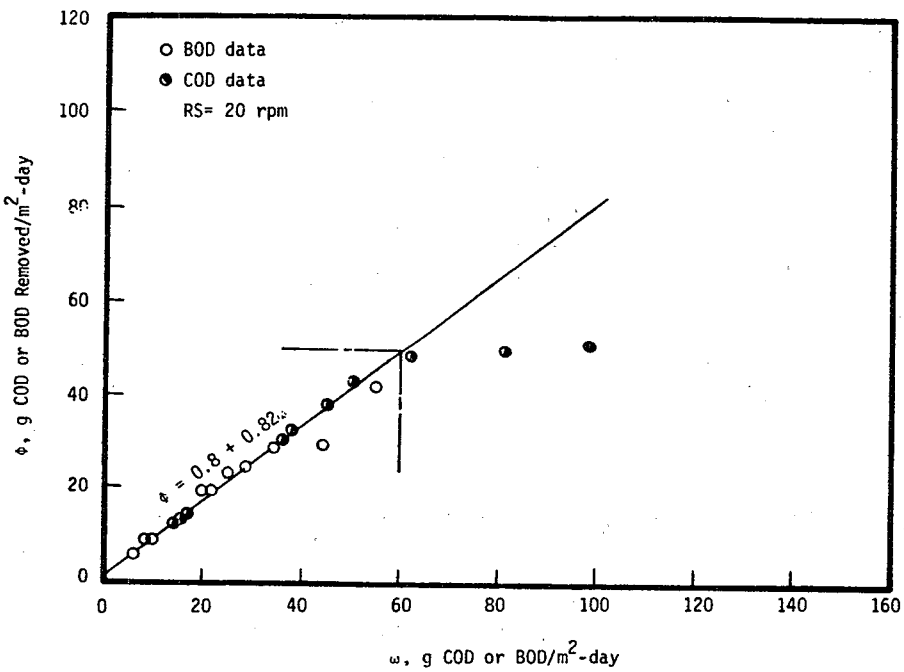


Figure 4. Realtionship Between Organic Loading and Substrate Removal Efficiency In Air RBC System

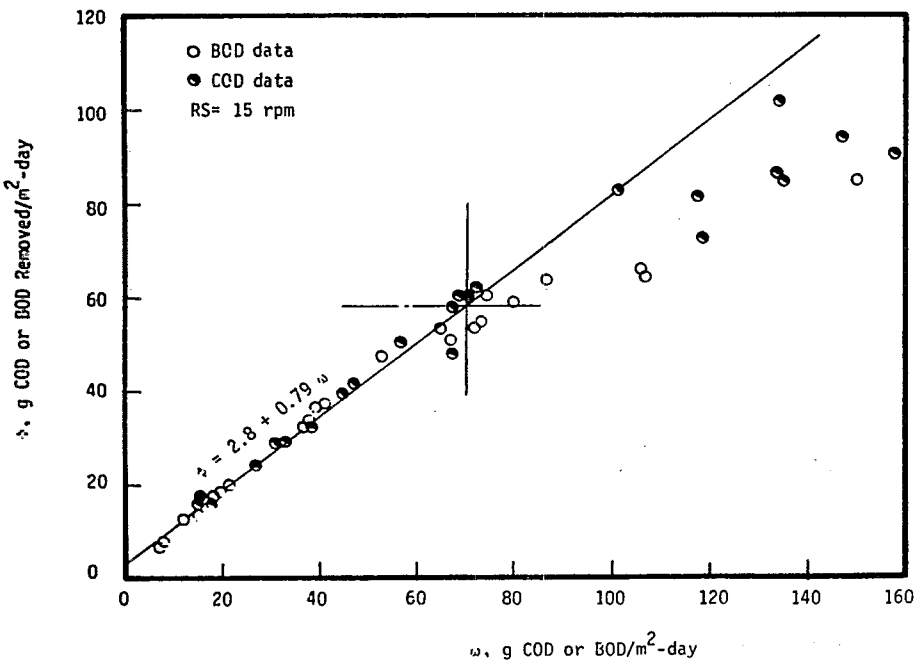


Figure 5. Relationship Between Organic Loading and Substrate Removal Efficiency In Oxygen RBC System

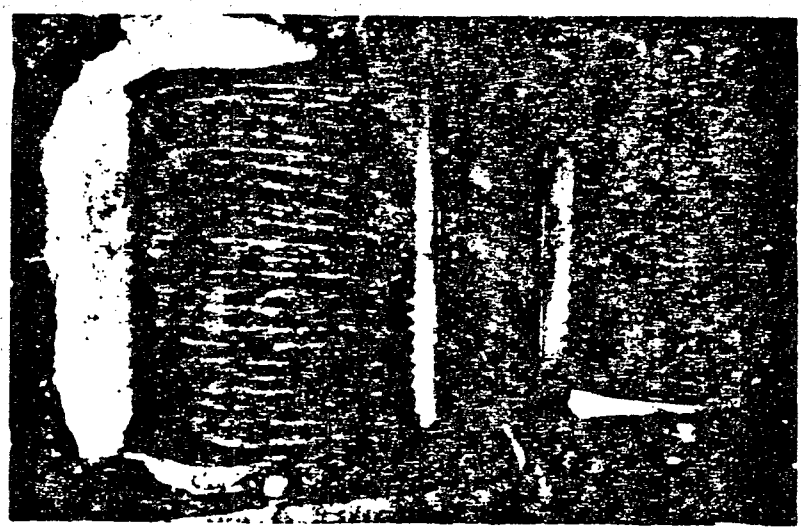


Figure 6. Foaming Problem in Air RBC System

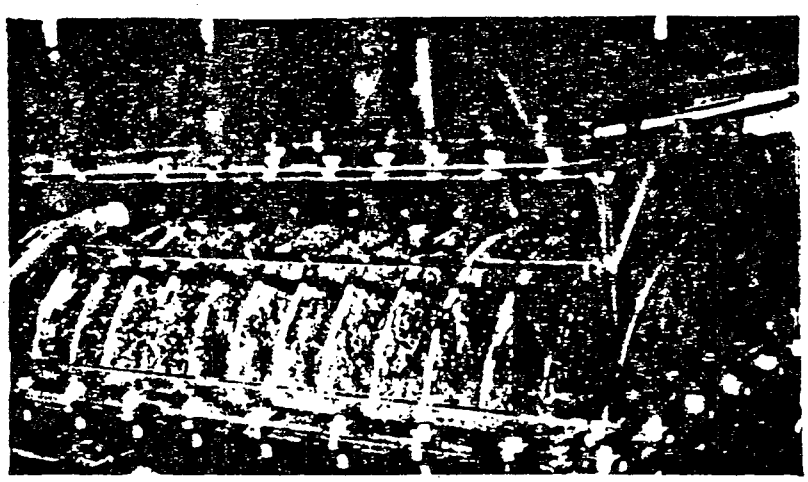


Figure 7. Disappearance of Foaming Problem In Pure Oxygen RBC System

RBC system. No biomass bridging was found. This result means that 4" disc spacing presently provided for the first RBC stage can be reduced to increase the disc number as well as the disc surface area. By doing so, the organic loading can also be reduced to increase the treatment efficiency.

Determination of β and K_s. Eq. 4 was used to calculate the values of β and K_s by plotting $(1/\lambda)$ against $(1/S^*)$ (see Figures 7 and 8). The β and K_s values obtained from the above calculations were summarized in Table 8 as follows:

Table 8. β and K_s Values

Parameter	Air RBC System		Oxygen RBC System	
	BOD Basis	COD Basis	BOD Basis	COD Basis
K_s, mg/l	151.4	890.4	10.9	432.0
β, mg/l-day	58.8	153.8	31.7	117.6

By substituting β and K_s values into Eq. 4, four RBC kinetic models are formed

Air RBC System:

$$\frac{A_w}{Q(S_o-S^*)} = 2.574\ (1/S^*) + 0.017 \quad \text{on BOD basis} \qquad \text{----------(7)}$$

$$\frac{A_w}{Q(S_o-S^*)} = 5.783\ (1/S^*) + 0.0065 \quad \text{on COD basis} \qquad \text{----------(8)}$$

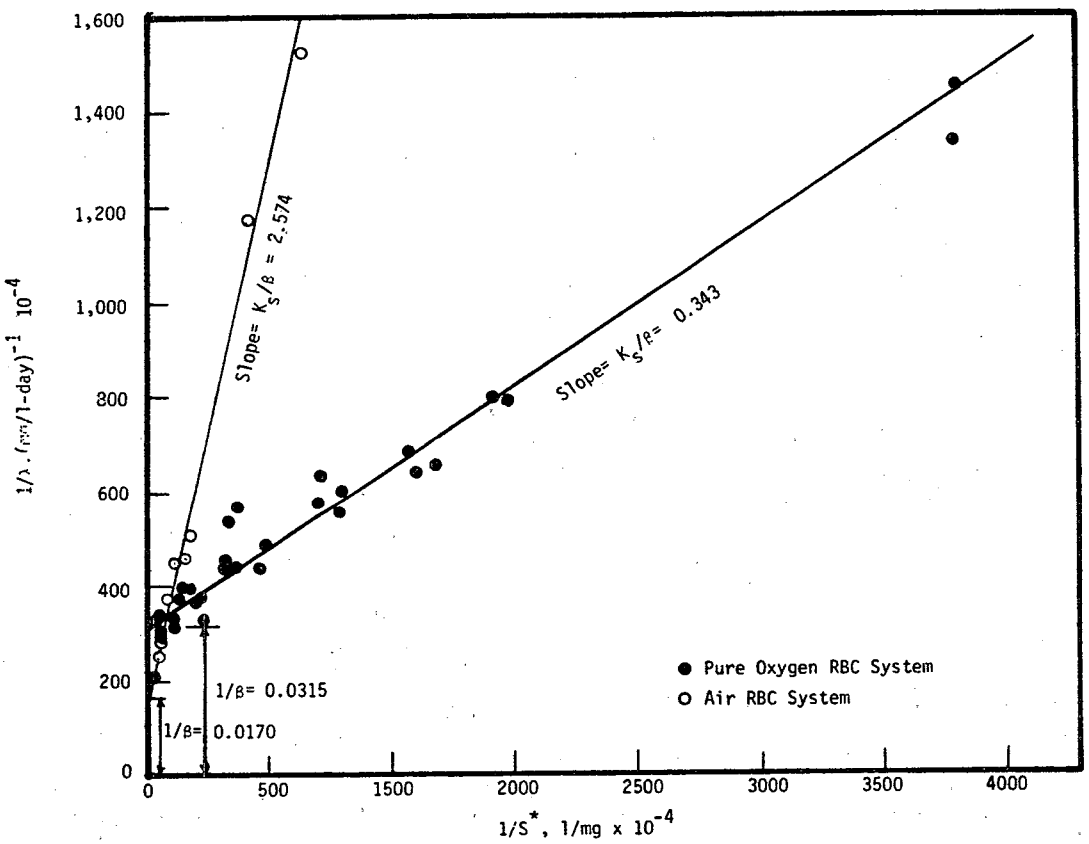


Figure 7. Determination of β and K_s On BOD Data Basis

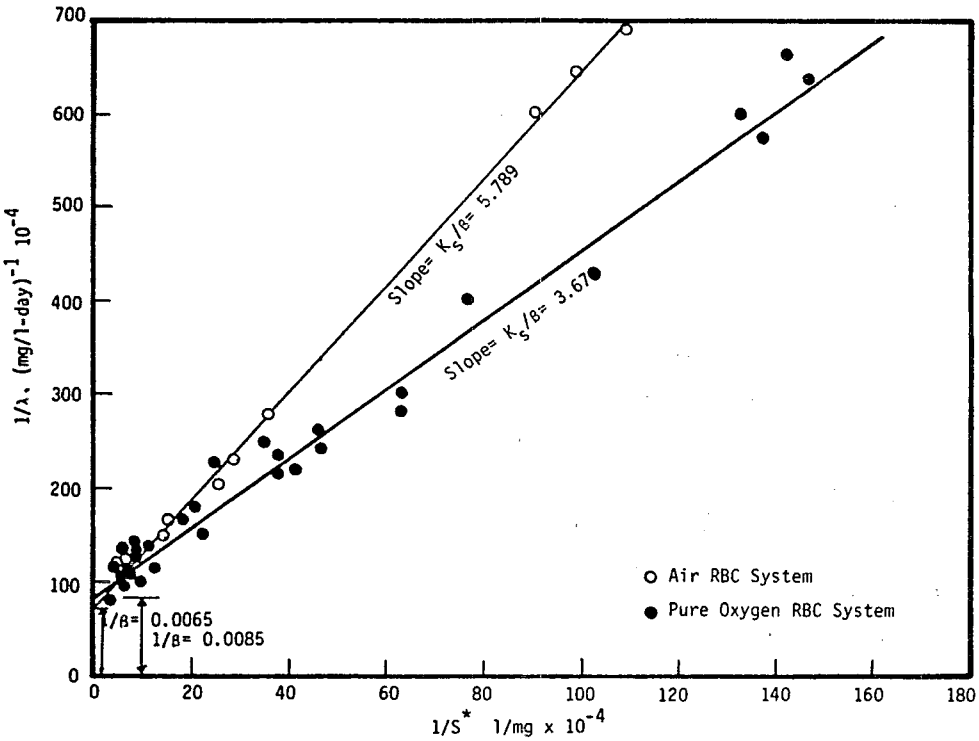


Figure 8. Determination of β and K_s on COD Data Basis

Oxygen RBC System:

$$\frac{A_W}{Q(S_o - S^*)} = 0.344\ (1/S^*) + 0.0315 \text{ on BOD basis} \qquad \text{----------(9)}$$

$$\frac{A_W}{Q(S_o - S^*)} = 3.673\ (1/S^*) + 0.0085 \text{ on COD basis} \qquad \text{----------(10)}$$

Conclusions

The following conclusions were formulated as a result of this study:

1. Proper pH control is essentially required for the treatment of this acidic wastewater. By adjusting the wastewater pH to nearly 9.0, the final effluent pH after passing through three RBC stages would be within the range of 7.0-7.6. The pH control was made by the addition of NaOH along with phosphate buffer.

2. In the pure oxygen RBC system, the disc rotational speed did not significantly affect the results of % COD or BOD removal, oxygen consumption, pH and DO level in each RBC stage. At the constant rotational speed of 15 rpm (12.3 m/min.), the RBC plant performance was closely related to the influent substrate concentration and organic loading as well.

3. Both air and pure oxygen RBC systems are capable of removing COD and BOD sufficiently. The COD and BOD removals were more than 90% when the organic loading in the conventional air system was below 28 g. BOD/m^2-day and 62.3 g. COD/m^2-day or it was less than 38 g. BOD/m^2-day and 110 g. COD/m^2-day in the pure oxygen system. From the present study, it was found that the latter system was possible to be operated at the organic loading 1.7 times higher than the former system if the efficiency of the RBC plant in terms of BOD removal equaled 80%. In general, the reduction of organic substrate was high in the first RBC stage when the organic loading was below 19.4 g. BOD/m^2-day in both systems. However, the subsequent

stages become important after the organic loading condition exceeding the above value.

4. The values of β and K_s in Eq. 4 are obatinable. It is apparent from the present study that the pure oxygen RBC system has lower β and K_s as compared to the conventional air RBC system on both BOD and COD data basis.

5. Two serious problems which may occur during the operation of RBC systems are: (a) foaming and (b)septic environment. The foaming problem started at the organic loading = 38 g. COD/m^2-day and the anaerobic suitation was developed at the loading equal or greater than 57 g. COD/m^2-day in the conventional air RBC system. Although no foaming problem was found in the pure oxygen RBC system, the septic condition occurred in the first stage at the organic loading = 95 g. COD/m^2-day. Poor sludge settling characteristics was resulted from the overgrowth of thread-like organisms under anaerobic condition.

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INHIBITION OF NITRIFICATION BY CHROMIUM IN A BIODISC SYSTEM

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INTRODUCTION

A series of investigations was undertaken to determine the acceptability of certain industrial wastes containing hexavalent chromium by a biodisc system providing both secondary treatment and biological nitrification.

The initial objectives of the study were: to determine the extent to which a biodisc system can tolerate chromium (VI) without losing efficiency in either BOD removal or in nitrification; to establish mechanisms of chromium removal and the benefits of staging, and to understand differences in short and long-term effects and steady state or shock load conditions.

These objectives were established to serve a number of purposes. The data gathered will assist municipalities in determining the quantity and characteristics of chromium containing waste that may be accepted without causing interference in operation. The data gathered will also assist design engineers in understanding and better defining the process reliability between an activated sludge and a biodisc system and finally the benefits of staging in an RBC.

Numerous studies on the effect of metals on biological nitrification have been reported in the literature in recent years, all of which have been confined to activated sludge(1)(2)(3). For this reason, this study on the effects of hexavalent chromium on the biodisc system was undertaken. In order to minimize differences due to other effects, a pilot plant was used.

EXPERIMENTAL METHODS AND APPARATUS

The shape and dimensions of the pilot system used are shown in Figure 1. Each biodisc tank was equipped with two sets (4 each) 2 ft. dia. disc media, a partition that separated the two adjoining stages, a drive with chain and speed controller. Two parallel systems were built for this experiment, each containing three, two stage units. Light weight concrete fillets, 3" x 3", coated with paraffin were placed at the bottom corners to prevent sludge accumulation and to improve the mixing pattern in the tank.

The standard substrate or feed solution was prepared such that the characteristics would closely simulate those of a typical municipal wastewater; BOD_5 at 200 mg/l, COD at 300 mg/l, total nitrogen at 20 mg/l. Dextrine was selected to be the major carbon source due to its slow biodegradation rate.

A stock solution of hexavalent chromium was prepared from $K_2Cr_2O_7$ and fed at a predetermined concentration either at the first stage or the fifth stage depending upon the purpose of the particular experiment. Typically, the slug loads of chromium were fed at the fifth stage to test short-term effect on nitrifying cultures. Long-term effects on the other hand, were studied by introducing chromium at the first stage.

A stock solution containing glucose was prepared and fed to test the effect of high organic loads on nitrifiers.

Table 1. Standard Feed Solution

Dextrin	150 mg/l	Ivory Soap	6.3 mg/l
Urea	42 mg/l	Consume Soup	2.1 ml/l
Na_2HPO_4	15.9 mg/l	Ann Arbor tap water	
$CaCl_2$	5.6 mg/l		
KCl	5.6 mg/l		
NaCl	12.1 mg/l		
$MgSO_4$	4.0 mg/l		

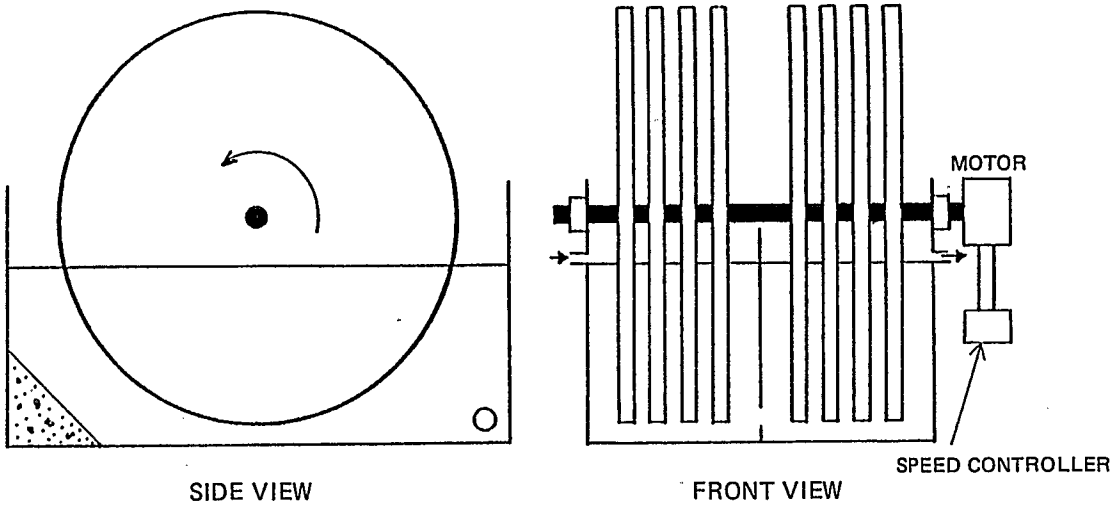

FIGURE 1. PILOT BIO-DISC SYSTEM

RESULTS AND DISCUSSIONS

A. Long Term Effect of Chromium

Three basis problems were present in this investigation: (1) the concentration at which the effects of a given metal ion are felt when fed continuously; (2) the concentration necessary to have a definite effect on a plant and the time required; (3) the fate of chromium in the system.

In all cases the two parallel biodisc systems were operating satisfactorily in both COD removal and nitrification before chromium was introduced. From earlier experiments it had been determined that the hydraulic loading rate of 0.65 gpd/S.F. was appropriate for full nitrification(4).

The effects of 3 concentrations (1, 3, and 10 mg/1) of chromium fed at the first stage were studied in a preliminary manner at a hydraulic loading rate of 0.65 gpd/S.F. The chromium was fed at 1.0 mg/1 approximately for two weeks, while at other concentrations less than a week.

The efficiency of nitrification was represented by its end product, nitrate, NO_3, since the intermediate product nitrite, NO_2^-, was negligible throughout the experiment. The nitrification at 1 mg/1 of chromium in Run 2 was slightly hampered at the second, third and fourth stages and yet nearly completed at the fifth stage. The sixth stage picked up the difference and completed nitrification, see Figure 2.

Effects of chromium at 3 and 10 mg/1 on nitrification were immediate and definitely inhibitory. The nitrate concentrations were reduced by 65 and 75 percent at each dosage in Runs 3 and 4, respectively.

On the basis of these preliminary findings, a long-term investigation of the chromium effect on nitrification began. The chromium concentrations chosen were 1 mg/1 in Run 5 and 2 mg/1 in Run 6. In addition, the hydraulic loading rate was doubled to 1.3 gpd/S.F. to expand the breadth of the investigation.

As shown in Figure 3, for a system receiving 1 mg/1 of chromium, the concentration of ammonia N continued to change for approximately a month before a relatively stable performance level was reached for the system.

Concentrations of other parameters such as COD, NO_2-N, and NO_3-N at this steady state are shown in Figure 4. The following observations were made:

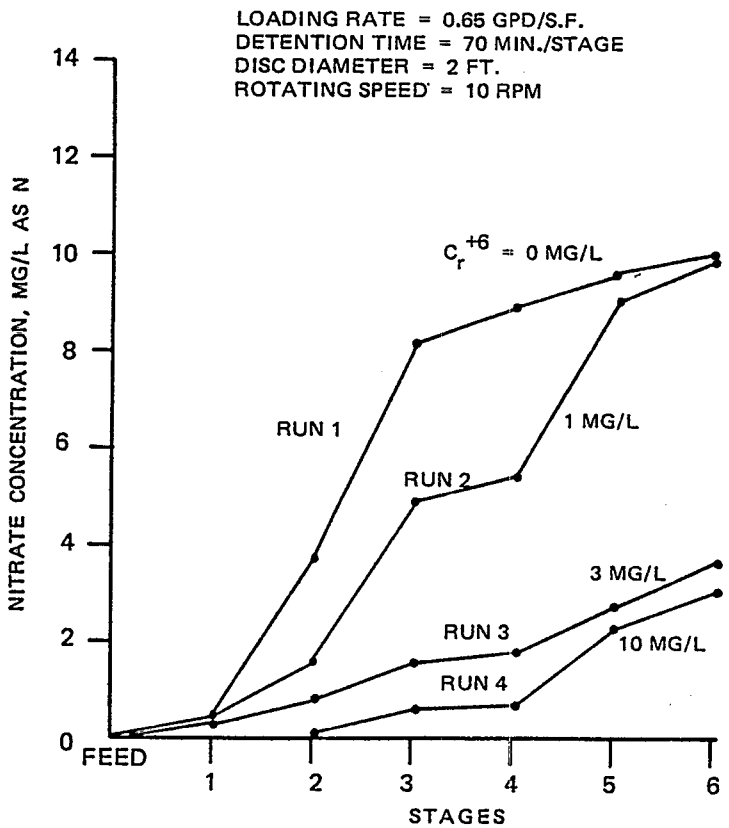

FIGURE 2. EFFECT OF CONTINUOUS CHROMIUM FEED ON NITRIFICATION

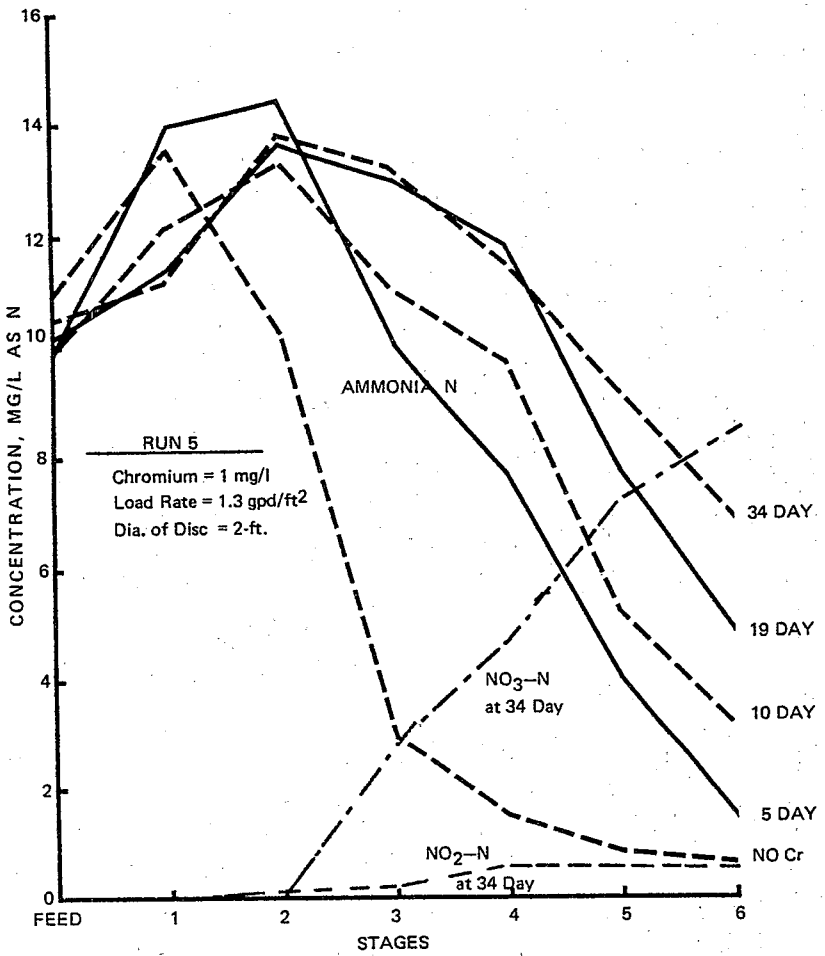

FIGURE 3. LONG TERM EFFECT OF CHROMIUM ON NITRIFICATION AT 1 MG/L

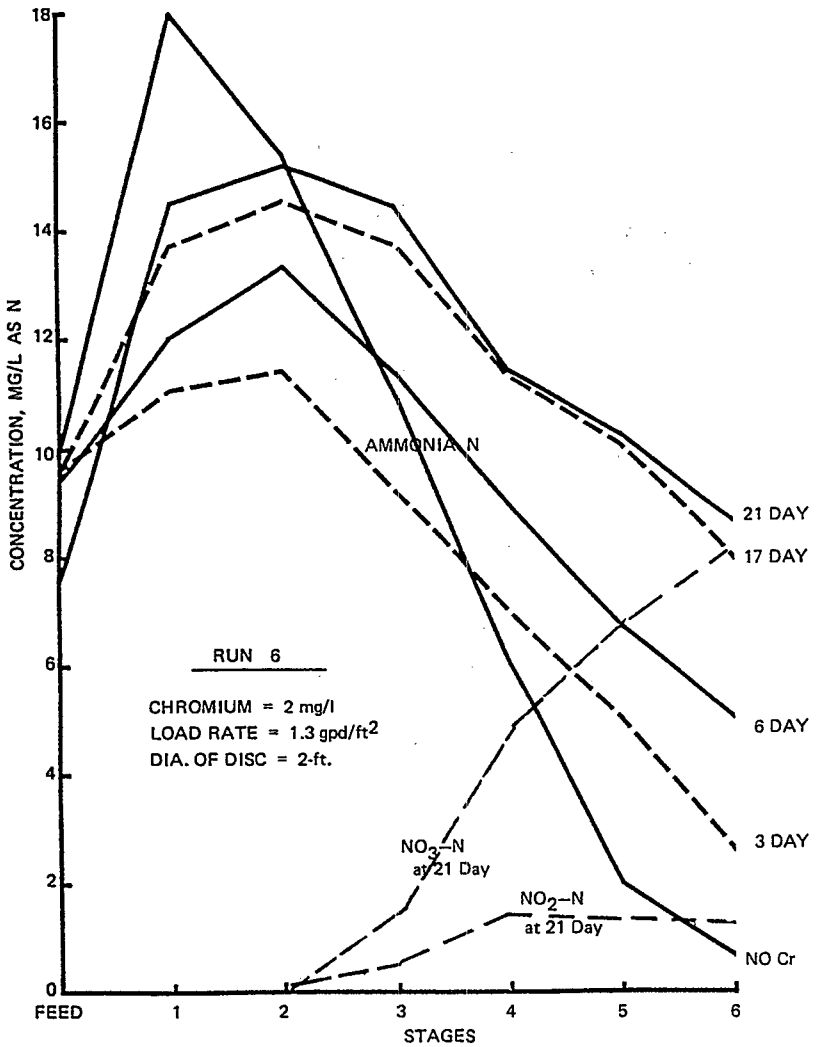

FIGURE 4. LONG TERM EFFECT OF CHROMIUM ON NITRIFICATION AT 2 MG/L

- Most of the COD was removed in the first two stages.
- Ammonification of organic nitrogen became complete at the third stage, compared to the first stage in Run 1.
- Nitrification of ammonia began at the third stage and continued in the subsequent stages
- No accumulation of nitrite was observed, indicating that the Nitrobacter group of organisms were not inhibited. Only Nitrosomonas was inhibited by chromium.
- Chromium concentrations in the liquid decreased rapidly in the first two stages (1.1 to 0.6 mg/1) and slowly in the subsequent stages (to 0.4 mg/1).
- The major mechanism of chromium removal from the liquid appeared to be adsorption to the biomass. The chromium content in the biomass layer closely approximated the COD profile throughout the treatment system. In the first stage for example, the chromium constituted approximately 2 per cent of biomass on a dry weight basis. In the following stages, the chromium content ranged between 0.6 and 1.0 percent.
- Staging definitely worked in favor of organisms in the later stages. While organisms initial stages were exposed to chromium, those in the later stages were not.

For the parallel system receiving 2 mg/1 of chromium, Run 6, similar observations were made, see Figures 4 and 5. The time required to reach a steady state operating condition however, appeared shortened; 21 days. The following observations were made:

- COD was being removed substantially in the initial two stages but slowly in the middle two stages.Overall removal was satisfactory.
- Ammonification of organic nitrogen also took three stages to complete.
- Nitrification began at the third stage and continued in the subsequent stages.
- Chromium concentration in the liquid decreased rapidly from 1.9 to 1.2 mg/1 in the first stage and remained between 1.2 and 1.1 mg/1 in the subsequent stages.

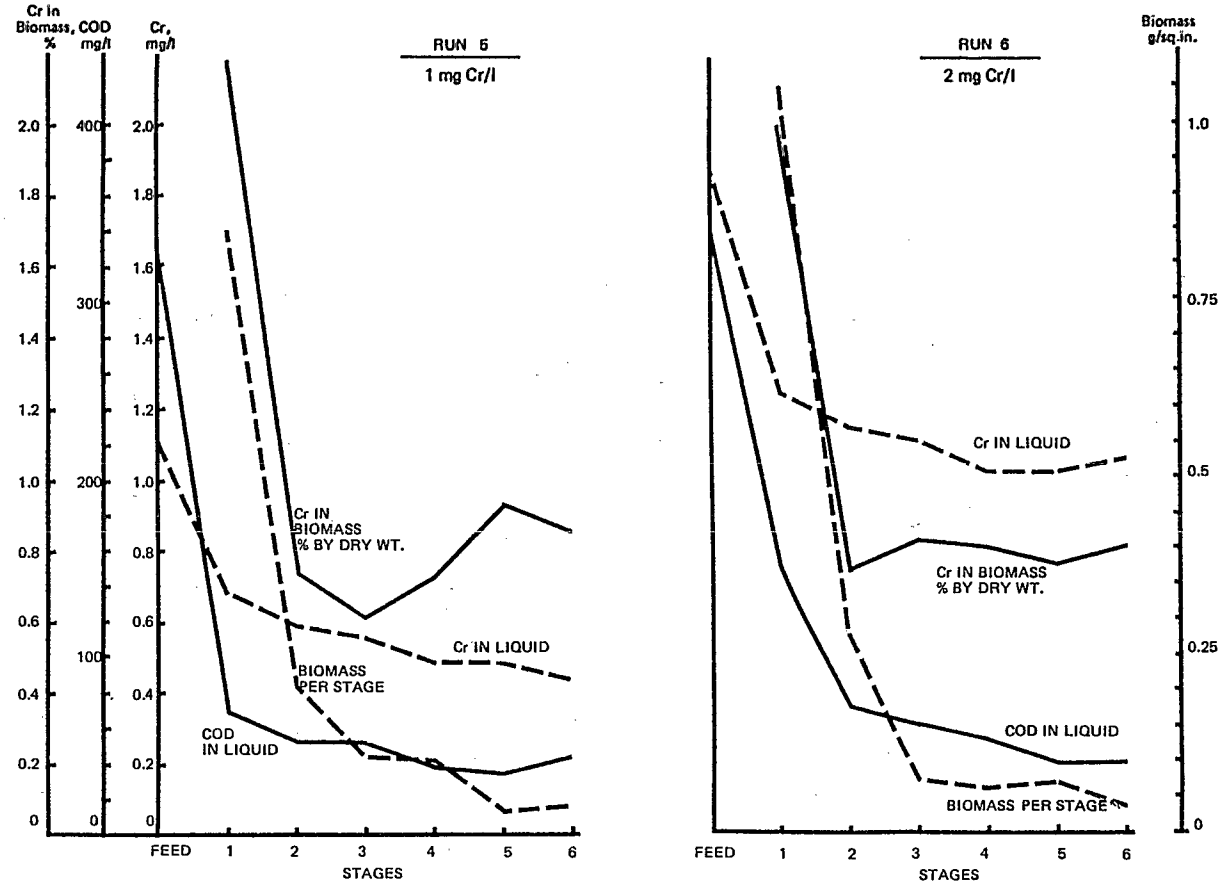

FIGURE 5. BIOMASS AND CHROMIUM PROFILES AT STEADY STATE

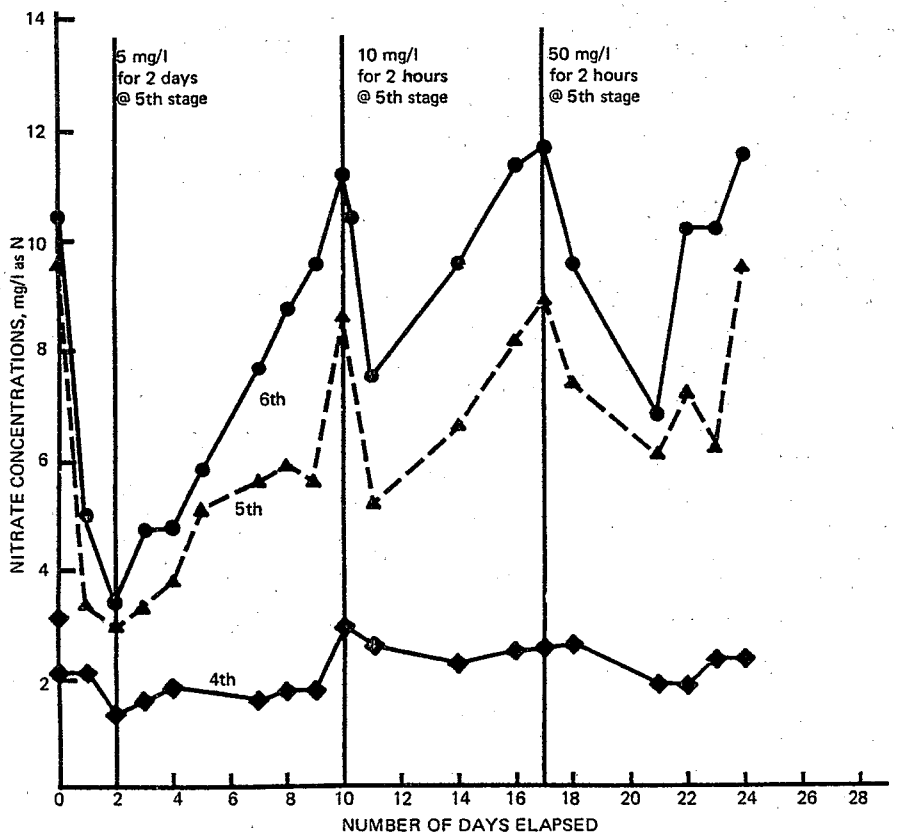

FIGURE 6. SHOCK RESPONSES ON NITRIFYING CULTURES

- The major mechanism of chromium removal is by adsorption to the biomass. The chromium content in the biomass was approximately 2 per cent in the first stage and 0.8 per cent on a dry weight basis in the last stage.
  Since the organic strength of the feed solution was the same between these two systems, the resultant biomass characteristics and quantity should be similar. It was then not surprising to observe that the adsorption of chromium by the biomass at steady state exhibited a similar mass relationship. Due to a higher concentration gradient in the more highly loaded system however, steady state appeared to have been reached sooner in the latter system (Run 6) than in the former where 1 mg/l of chromium (Run 5) was used.
- Among the organisms involved, it may be concluded again that the more sensitive of the two groups of nitrifiers is Nitrosomonas. Nitrobacter appeared to be less sensitive to chromium and thus to oxidize nitrite without much accumulation of nitrite.
- Benefits of staging were again observed in that inhibition took effect by stage. When a load came, the late stages would not be affected, even if the initial stages were adversely affected.

From the data presented in the section, one could summarize that:

Hexavalent chromium could be adsorbed to the biomass upon contact up to its adsorptive capacity (2 per cent for heterotrophic organisms and less for a mixture of heterotrophic and autotrophic organisms). This is a higher level than 0.8 percent as reported on activated sludge(3). Since the biomass density of biofilm is higher than activated sludge floc, it appears that the adsorptive capacity improves in a biodisc system.

Inhibition was controlled more by what was in the biomass layer than in the bulk liquid. Even though the bulk liquid carried chromium at or above 0.6 mg/l, the effect was not immediate.

Staging helped the process as a whole. Benefits could be achieved in two ways: While an upstream stage received and removed chromium, the downstream stages were spared from the inhibitory impact or received a minimal quantity, all of which increased the overall process reliability.

Since the chromium concentration in the initial stages would be higher than the case where staging was not practiced, the rate of adsorption to the biofilm would be faster. Under a slug load, the chromium would concentrate in one stage and upon sloughing, could be removed from the system in an efficient manner.

Nitrosomonas showed a greater degree of inhibition by chromium than Nitrobacter, which was consistent with observations elsewhere(2).

B. Short Term Effect of Chromium

The problems generated in this phase of the study were to investigate: (1) that concentration which would produce definite inhibition and the resulting reaction time.
(2) the rate of recovery, if any, once the slug is passed.
(3) acclimation effect under repeated slug doses.
(4) the impacts of heterotrophic organisms on nitrifiers.

In this study concerned with the effect of slug doses of hexavalent chromium on the nitrification process, concentrations of chromium at 5 mg/l for 2 days (Run 7), 10 mg/l for 2 hours (Run 8) and 50 mg/l for 2 hours (Run 9) were used in succession. Approximately a week of recovery was allowed between runs. These concentrations were fed at the fifth stage since at this point the system was supporting full nitrification. The efficiency as measured by the reaction product, or nitrate, NO_3^-, dropped immediately as shown in Figure 6. As soon as the chromium feed was removed from the system however, the nitrification was shown to resume and within approximately a week had returned to its original level. The degree of inhibition appeared to be inversely proportional to the mass loading. It was postulated that this rapid resumption of nitrification was made possible by three contributing factors: (1) the adsorption of chromium to the biomass layer which appeared to have caused an immediate inhibition of the existing culture and yet remained within the biomass layer, where its presence caused no further inhibition of new growth. The desorption of chromium did not appear to be significant; (2) continuous seeding from the fourth stage which contained nitrifying cultures; (3) acclimation of cultures to chromium was a definite condition which minimized the degree of inhibition in Run 8 (with 50 mg/l chromium). In this case inhibition was not as severe as in Run 7 where 10 mg/l chromium was fed.

C. Impact of Presence of Heterotrophs on Nitrification

In an effort to investigate nitrifying organisms competetiveness in the presence of heterotrophic organisms, 50 mg/l of glucose was fed to the fifth stage for three days in a similar manner as chromium had been previously introduced.

Results are shown in Figure 7 that indicate with the growth of heterotrophs on the surface, nitrification efficiency starts to decline. Once the glucose feed was removed, there is no immediate increase in nitrification activity.

It appears that nitrifying cultures are outgrown by heterotrophs on the same surface, thereby they become buried among the heterotrophs. With the increased oxygen demand exerted by heterotrophs, the dissolved oxygen concentration in the bulk liquid decreased to 1.0 mg/l and as a result was limited for nitrifiers. This data confirms a theory that under a d.o. suppressed environment, nitrifiers cannot function properly and therefore a nitrogenous oxygen demand cannot exist.

SUMMARY AND CONCLUSIONS

Chromium may enter municipal waste treatment plants in many different ways. Perhaps most frequently it occurs in plating waste, although it may have its source in tanning operations, in waters given corrosion inhibition treatment with chromate, or in aluminum-anodizing wastes.

Concentrations of hexavalent chromium up to 10 mg/l were fed up to four months on a continuous basis to two six-stage biodisc systems treating synthetic sewage, with concentrations of 300 mg/l of COD and 20 mg/l of nitrogen. Initial stages of the discs were supporting heterotrophic organisms, while the biological nitrification was achieved by autotrophs in the later stages.

Concentrations of hexavalent chromium up to 50 mg/l were also fed as slugs to the biodisc system to test the effects of such doses.

In addition, the effects of the presence of heterotrophic organisms on nitrifiers was examined by feeding glucose directly into a nitrifying stage.

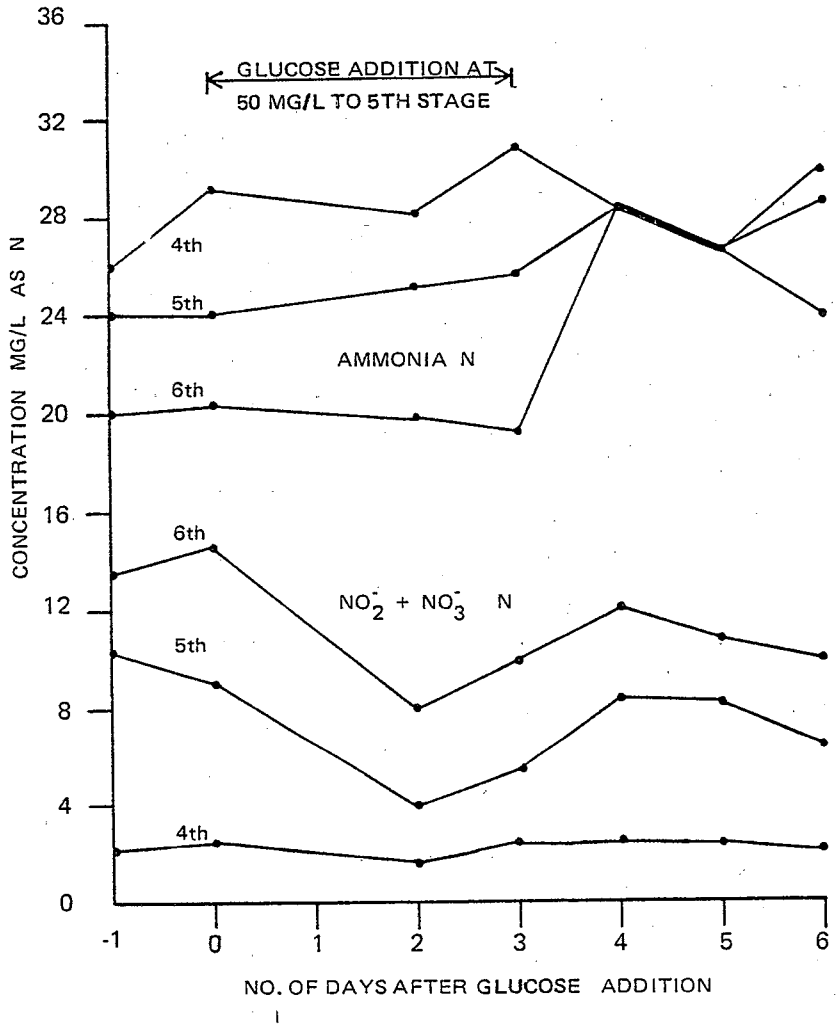

FIGURE 7. EFFECT OF GLUCOSE ADDITION ON NITRIFICATION

From the observations made in this investigation, the following conclusions may be drawn.

On a continuous basis, 1 mg/l of hexavalent chromium exhibited a consistently inhibitory effect on nitrifying cultures.

The major mechanism of hexavalent chromium inhibition to microorganisms appeared to be by adsorption. When chromium is adsorbed to a discrete active microorganism, the inhibitory effect sets in quickly. Chromium also may be adsorbed to non-active biological film or flocs, thereby becoming immobilized. The degree of adsorption or affinity however, depends on the characteristics of the film or flocs in the system. For example, heterotrophic cultures established in the initial stages, oxidizing primarily carbonaceous materials exhibited chromium retention of approximately 2 per cent, while the mixture of nitrifying cultures containing a limited quantity of heterotrophic organisms established in the later stages, exhibited a concentration of approximately 0.8 per cent on a dry weight basis.

As previously indicated data gathered in this investigation showed that the rates of adsorption and the resulting inhibition to microorganism were fast.

In a matter of hours the inhibition at a stage was well defined. When the adsorptive capacity of the entire biomass film in a stage was reached, the impact was shown by reduced COD removal for heterotrophs and reduced nitrate generation by nitrifiers.

Data also indicated that between two major groups of nitrifiers, Nitrosomonas was more sensitive to chromium and thus ammonia oxidation was decreased, while the nitrite oxidation to nitrate remained unchanged.

Data further indicated that staging offered advantages in process reliability and also in isolation and efficient removal of affected sludge from the system.

The impact of slug loads of chromium on nitrification was tested at concentrations up to 50 mg/l. Since the biofilm at the fifth stage composed mostly of nitrifying cultures, the chromium addition at that stage resulted in immediate reduction in nitrification, even at 5 mg/l. The complete recovery of the system however was rapid, taking place within approximately a week, indicating that the chromium retained in the film and flocs did not adversely affect the newly developing cells being established. Desorption of these chromium compounds appeared not to have generated a problem. The RBC system also showed a benefit due to continuous seeding from the upstream stage.

An advantage due to previous exposure to chromium was also shown. Acclimation was evidenced from the studies of slug doses.

The fact was likewise confirmed that when heterotrophs start to grow on the same support surface, nitrifiers cannot compete for either surface or oxygen and thus become overgrown.

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ACKNOWLEDGEMENTS

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PART IX: INDUSTRIAL WASTEWATER TREATMENT

SCALE-UP AND PROCESS ANALYSIS TECHNIQUES FOR PLASTIC MEDIA TRICKLING FILTRATION

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ABSTRACT

Reaction models for bio-oxidation using a sheet flow reactor with a fixed biological film similar to that used in plastic media trickling filtration are developed. The models utilize plug flow hydraulics and accept various descriptions of BOD removal kinetics including: zero order, first order, retardant and concentration dependent mechanisms. Hydraulics, kinetics and film geometry are individually incorporated.

System model equations are arranged into linear expressions which allow graphical determination of model applicability and rate constants from plots of experimental data.

The design and operating characteristics of laboratory simulation equipment are presented. Simulation equipment consists of continuously fed inclined planes with effluent sedimentation and recirculation. Effluents can be fed from and returned to coldroom storage. Heating tapes control temperature above ambient levels. System location in a cold-room provides temperature control at below ambient levels.

Applicability of the models is verified using operating data from laboratory and full scale studies of a number of effluents including: municipal sewage, whey wastewater, kraft

mill effluent, sulfite mill discharges, hard-board mill effluent, yeast fermentation effluents, pharmaceutical discharges and meat processing wastes. Additional verification is presented from literature data utilizing a glucose substrate. Scale-up calculations are developed to utilize laboratory data to determine full scale performance for various packing geometries. Design calculations are also presented including determination of the influence of recirculation.

INTRODUCTION

The trickling filter is comprised of a bed of media on which biological film growths develop. Removal of BOD is obtained by aerobic processes at the film surface and by anaerobic processes within the film interior. Modifications to the classic rock trickling filter introduced plastic geometric media to obtain increased surface area and porosity. Current practice generally employs a lattice type structure of vertically oriented media which induces a sheet flow regimen.

Reactor operation is such that laboratory scale simulation can be used for rate constant determination and in parallel operation with pilot plant equipment to reduce data collection requirements and extend interpretation of pilot scale results. Simulation equipment utilizes an inclined plane of up to 18' in length operating with thermal regulators and evaporator control systems.

The theory of BOD removal by trickling filter slime over a reaction surface similar to that of an inclined plane has not been formulated in the existing engineering literature. Full scale design equations in current use have been frequently developed by emperical methods or by analogy to formulations used to describe BOD exertion in general. The work presented herein describes development and verification of reaction models suitable both for laboratory studies and for scale-up to prototype conditions.

Consideration is given to the analysis of laboratory scale data, the interpretation of pilot and full scale results and the design of prototype systems for each reaction model. The theoretical development is presented in the following sequence:

- o General reaction model for inclined planes
- o Specific models for alternative bio-kinetic rate processes
- o Influence of recirculation and temperature

- Scale-up techniques for full scale conditions
- Design Procedures

General Reaction Model

A general model for BOD removal on an inclined plane surface is obtained by the solution of a material balance statement in which the hydraulics of liquid flow and the kinetics of biological reaction have been separately included. A schematic of an inclined plane system is shown in Figure 1.

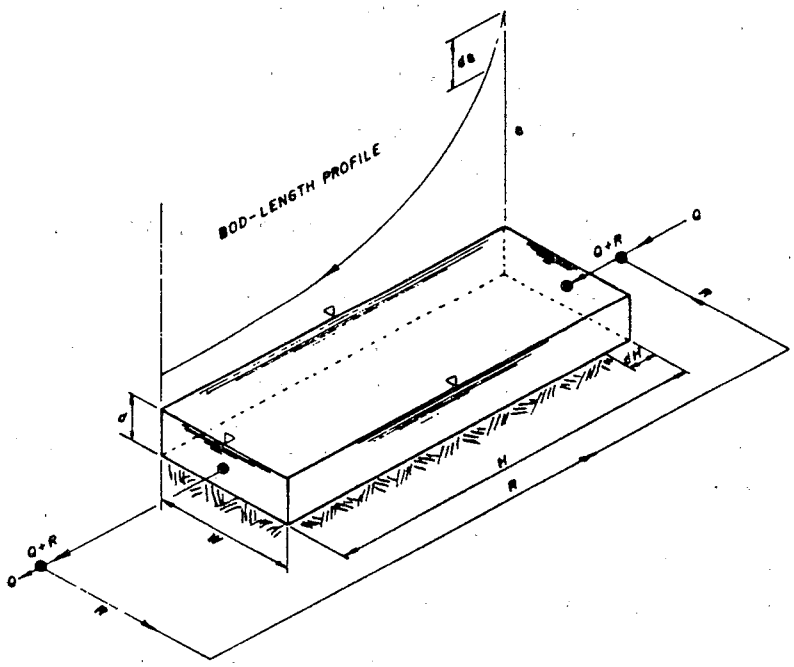

Figure 1 Schematic Representation, BOD Removal Over Slimed Surface

The material balance statement is written thus:

INPUT - OUPUT - REMOVAL = ACCUMULATION

INPUT and OUTPUT terms are self explanatory. The REMOVAL term is defined by the geometry of the reactor and the kinetics of the biological reaction. The ACCUMULATION terms accounts for

the change in the quantity of BOD stored in the reactor volume. For both inclined plane surfaces and trickling filters this storage is negligible. The terms of the material balance statement are mathematically defined as follows:

$$\text{INPUT} = (Q+R)S_a$$

$$\text{OUTPUT} = (Q+R)S_e$$

$$\text{REMOVAL} = K_r V_r$$

$$\text{ACCUMULATION} = 0$$

where:

Q = Untreated flow (gal min^{-1})

R = Recirculation flow (gal min^{-1})

S_a= BOD concentration as applied to reactor (mg/l)

S_e= BOD effluent concentration-mg/l

V_r=Volume of reactor (gal)

K_r= A generalized BOD removal rate constant for the reactor volume (mg/l min^{-1})

The reactor is examined over a differential element of plane height (dH) to define reactor volume:

$$dV_r = (d)\ (W)\ (dH) \qquad (1)$$

The material balance for BOD input and removal is then stated as follows:

$$(dS)\ (Q+R) = (Kr)\ (d)\ (W)\ (dH)$$

which when rearranged into a differential statement becomes:

$$-\frac{dS}{dH} = \frac{K_r(d)\ (W)}{(Q + R)} \qquad (2)$$

The term, $(Q + R)/W$ is conveniently grouped as a hydraulic loading per unit of plane width (U) as follows:

$$-\frac{dS}{dH} = \frac{(K_r)\ (d)}{(U)} \qquad (3)$$

Equation (3) then is the general statement for BOD removal over a slimed surface.

First Order Reaction Kinetics

At this juncture biological reaction kinetics may be introduced to develop a specific reaction model, e.g., first order, retardant etc. A first order reaction model has been found in practice to apply most frequently to trickling filter performance.

The generalized rate constant (K_r) for a first order reaction is expressed as the product of a specific velocity constant and the amount of substrate present. The relationship in generalized terms is:

$$K_r = (k)\ (S) \qquad (4)$$

The units used for k and S are dependent on the way in which the reaction is described e.g. in activated sludge work S has the units of concentration and k is expressed as a substrate concentration change, per unit of substrate and organisms present, per unit of time. K_r then has the net units of time^{-1}.

In fixed film reactors the specific velocity constant is expressed in terms of the unit weight of substrate removed, per unit weight of film present, per unit of time.

$$k = \frac{dM_s}{dt\ M_f}$$

where:

dMs = substrate weight change
dt = time change
M_f = film weight present

By expressing substrate weight in terms of concentration, i.e. $dM_s = V_r dS$, k then becomes:

$$k = \frac{V_r dS}{d_t M_f} = \left(\frac{dS}{dt}\right) \left(\frac{1}{\frac{M_f}{V_r}}\right) \qquad (5)$$

M_f/V_r can then be termed the equivalent concentration of organisms in the elemental reactor volume being analyzed. For the inclined plane:

$$\frac{M_f}{V_r} = \left(\frac{d_f}{d}\right) \left(g_f\right)$$

where:

d_f = depth of fixed film
d = depth of flow q
g_f = specific gravity of film

These definitions then provide the expression for the generalized reaction rate as follows:

$$K_r = (k') (S) \qquad (6)$$

wherein

$$k' = (k) (d_f) (g_f)$$

The effective slime concentration provides for an active weight of organisms per unit of area and thereby implies that a thickness of film is operating to effect BOD metabolism. This is indeed the case and has been experimentally verified by Hoehn and Roy (1). Figure 2 presents their data for COD removal vs. film thickness.

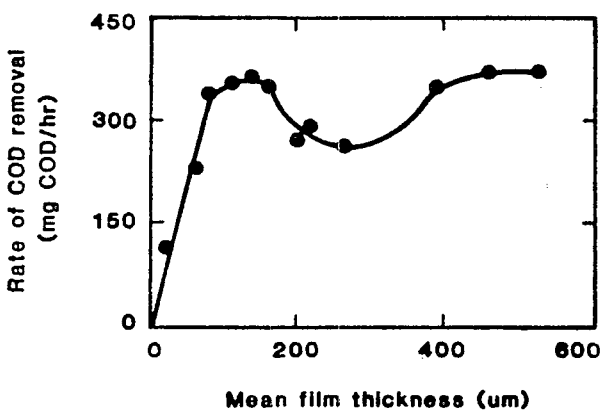

Figure 2 Rate of COD Removal as a Function of Film Thickness (1)

The description of a first order reaction may then be completed as follows:

$$-\frac{dS}{dH} = \frac{k'S}{U} \qquad (7)$$

It is worthwhile to note that the derivation of K_r proceded in such a way as to eliminate flow depth (d) from the final expression. This in turn results in an exponent of 1.0 for the flow term (U). If depth remained in the final equation then the fact that depth over an unslimed plane varies with $(U)^{-1/3}$ would result in an exponent of 2/3 for the flow term U. The elimination of depth also tacity implies that the reaction at the slime/water interface controls rather than the transport of substrate through the liquid depth to reach the interface. Equation (7) thus describes a reaction controlled model. The description of the active organism concentration(M_f/V_r) interms of a ratio of slime specific weight divided by liquid depth also results in an averaging process for the amount of biomass effectively operating to remove and metabolize BOD. The definition of the amount of biomass representedthe first half of the analysis. The balance of the (K_r) breakdown used kinetic expressions which are independent of organism concentration i.e. $K_r \sim k'S$. The effect of this two part definition is to effectively insert a linear averaging technique into the overall rate constant K_r.
Thus, any other BOD removal and metabolism mechanism which can be reasonably described by a linear average of a combination of liquid depth and bio-mass amount will be approximated by equation (7). This may help to explain why the first order formulation has been found to apply to performance data over a wide range of process conditions. Integration of expression (7) provides the basic analytical relationship for BOD removal over an inclined surface:

$$\frac{S_e}{S_a} = e^{-k'H/U} \qquad (8)$$

In the application of equation (8) to laboratory analysis an incline of up to 45^o from the horizontal may be used without concern for the non-formulated depth effects caused by angle of inclination.

Incorporation of Recirculation and Temperature

The ratio $\frac{S_e}{S_a}$ describes the change in BOD as applied to

the film and, therefore, implicitly includes the effects of recirculation. For design purposes BOD changes relative to the undiluted wastewater are required. The change in reactor BOD's is related to the undiluted wastewater by a material balance around the plane as follows:

$$\frac{S_a}{S_e} = \frac{1 + r\ (1-E)}{(1+r)\ (1-E)} = f \qquad (9)$$

where:

r = recirculation ratio R/Q

E = BOD removal efficiency based on raw wastewater

The factor (f) is introduced for topographical simplicity.

The effect of temperature on reaction rate is introduced using the Arrhenius relationship.

$$k'_t = k'_{20}\theta^{\Delta T} \qquad (10)$$

where:

k' = reaction constant at temperature t

k'_{20} = reaction constant at standard temperature, 20°C

Δt = reaction temperature differential °C-20

θ = constant, usually taken as 1.035

ANALYSIS OF LABORATORY PLANE PERFORMANCE

The analysis equation for first order performance is completed as follows:

$$f = e^{k'_{20}\theta^{\Delta t}\ H/U} \qquad (11)$$

A graphical solution to equation (11) is obtained by taking double logarithms as follows:

$$\log\left[\frac{H\theta^{\Delta T}}{2.3\log\ (f)}\right] = \log U + \log(1/k'_{20}) \qquad (12)$$

A plot of data on log paper will provide a linear correlation with slope equal to 1.0 and an intercept of log $(1/k'_{20^o})$ at

U = 1.0.

It is important to note that the variable H, height of plane, is portrayed as having a linear effect on plane performance. This is so in that laboratory planes are short and receive a controlled, uniform discharge. Both of these characteristics support a linear relationship between height, flow pathway, film presence and thickness etc. This condition does not necessarily apply to full scale towers.

An application of first order kinetics to inclined plane treatment of whey plus sewage, Quirk (2), is shown on Figure 3. Test results were obtained using plane lengths of 9 and 18 ft. operated at hydraulic loadings from 0.02 to 0.15 gpm/ft. Recycle ratios ranged from 0. to 6.0. Film areas of 0.375 and 0.750 SF provided the data shown operating over a range of organic loadings of 0.08 to 0.95 lbs BOD/day/SF Slime. An hydraulic coefficient of $n = 1.0$ and $k_{20^o} = 1.8 \times 10^{-3}$ gpm/SF were determined. Shultze (3) also experimented with a mixture of whey plus sewage using vertical meshed screens 3' x 6' in dimension as a fixed film reactor. His data correlated using a first order reaction as shown in Figure 4. Hydraulic effects produced an $n = 1.0$ with a reaction rate constant of $k_{20^o} = 2.5 \times 10^{-3}$ gpm/SF. Hydraulic loadings ranged from 0.08 to 0.45 gpm/SF and were about three times the hydraulic loadings used by Quirk (2).

Additional Kinetic Models

Kinetic models other than a first order reaction have been employed to describe biological oxidation. While less popular than the first order assumption, these additional models have been found to correlate bio-oxidation data in a successful manner. Because of a cumbersome mathematical structure when used for fixed film reactors, these models have been applied primarily to fluid bed reactors such as activated sludge, aerated stabilization basins, etc.

The relationships for laboratory planes have been extended to include:

1. Linear and exponentially retardant reactions
2. Concentration dependent reaction
3. Zero order or constant rate reaction

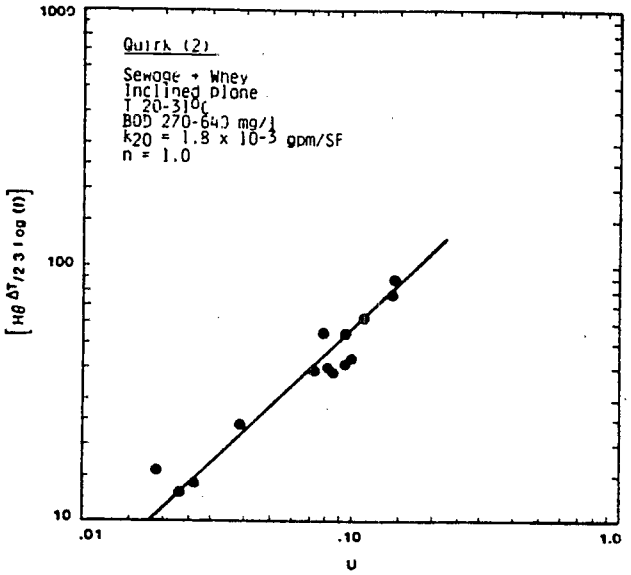

Figure 3 Whey and Sewage Performance on Inclined Plane (2)

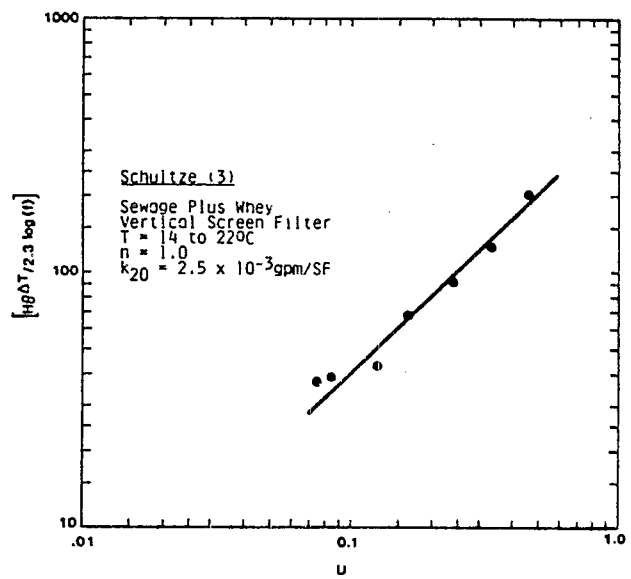

Figure 4 Whey and Sewage Performance on Vertical Screen Filter (3)

Retardant Reaction Models

In a simple retardant reaction the rate of BOD removal per unit weight of organism is proportional not only to the BOD concentration remaining but also to the fraction of BOD remaining. Rates of removal decrease, or retard, rapidly as high efficiencies of removal are approached and the relatively assimilable organics havebeen metabolized. The kinetic statement is written as:

$$K_r = (k')(M_f/V_r)(S)\ (S/S_a) \qquad (13)$$

After integration the completed equation for slimed plane analysis is written as:

$$f = 1 + \frac{(k)' \cdot (H)}{U} \qquad (14)$$

An exponantially retardant reaction can be described by the relationship:

$$(f) = \left[1 + \frac{Nk'H}{U}\right]^{1/N} \qquad (15)$$

Concentration Dependent Model

A concentration dependent reaction assumes a linear reaction of all substrate elements with the percent remaining as the expression of retardancy. Unlike the retardant reaction, the concentration dependent formulation does not relate removal rate to concentration remaining but rather to a maximum rate which would occur at zero removal. The equation for a slimed plane operating with concentration dependency becomes:

$$f = e^{k'H/USa} \qquad (16)$$

The combined parameters of H, U, and S_a represent the organic loading or F/M ratio per unit area of film. Thus:

$$f = e^{k'/F/M} \qquad (17)$$

a linear plot is obtained by relating log(f) to $k'/(F/M)$.

Inclined plane studies using glucose were found to be correlated by a concentration dependent reaction. The Maier(4) data is correlated by a concentration dependent model as shown on Figure (5). The separation of the 37.3°C data is apparent while the balance of the results fit a single model. The reaction rate for these glucose experiments was found from the

correlation slope at $k'_{20} = 0.50 \times 10^{-3}$ lbs/SF/Day. The existence of a vertical intercept a 1/(F/M) = 0 is not predicted by theory and appears to represent the effects of reduced hydraulic rates and increased glucose concentrations used to attain high F/M values.

A similar effect is evidenced by the data of Oleszkiewicz and Eckenfelder (5). This inclined plane data correlated using a concentration dependent reaction as shown in Figure (6). The wide variation in feed BOD concentration and recycle ratios used in these studies were all encompased by a single model. The phenomenon of a vertical intercept is again evidenced.

Zero Order Reaction

In a zero order reaction the rate of BOD removal per unit weight of orgamism is constant and is not effected by substrate concentration, degree of removal etc. The relationship is:

$$\frac{1}{f} = 1 - k'/_{(F/M)} \qquad (18)$$

Figure (7) illustrates the zero order reaction obtained on effluents from an Insulation Board mill operating on a wood and mineral raw material, Quirk (6). Inclined planes of 8 and 10 feet were used and operated over a range of 23 to 30°C. Raw wastewater BOD ranged from 1000 to 6500 mg/l. Recycle ratios of up to 20/1 were used. A k_{20} value of 5×10^{-3} lbs/SF/day was obtained for the highly spread data. A second illustration of the zero order reaction was obtained in a study of yeast fermentation effluents, Quirk (7), as shown in Figure 8. A reaction rate coefficient of 9.0×10^{-3} lb/SF/day at 20°C was obtained. A comparison between this data and subsequent pilot plant operation will be presented below.

SCALE-UP RELATIONSHIPS

Conversions to full scale tower conditions is made by adjusting plane performance for the following:

1. Hydraulic loading at full tower height.

2. Film thickness anticipated.

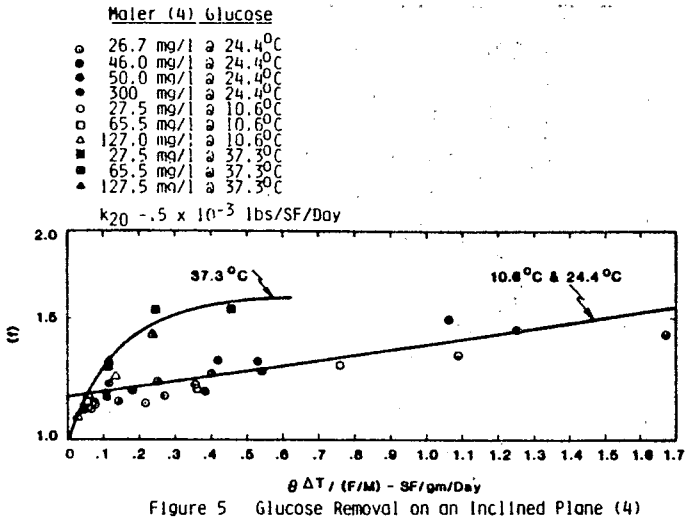

Figure 5 Glucose Removal on an Inclined Plane (4)

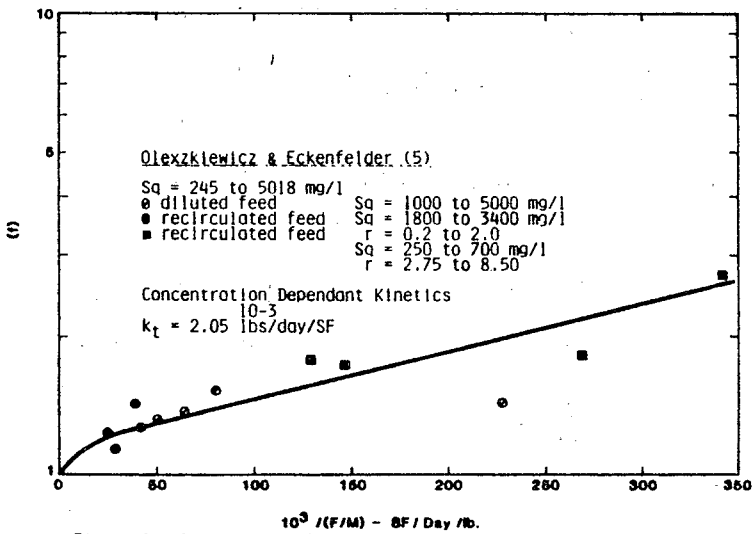

Figure 6 Pharmaceutical Waste Removal on an Inclined Plane (5)

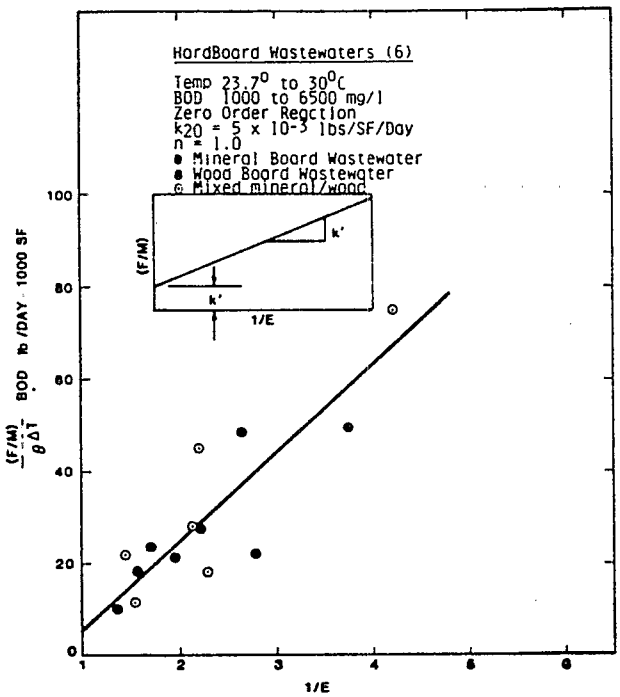

Figure 7 Hardboard Wastewaters Treated on an Inclined Plane (6)

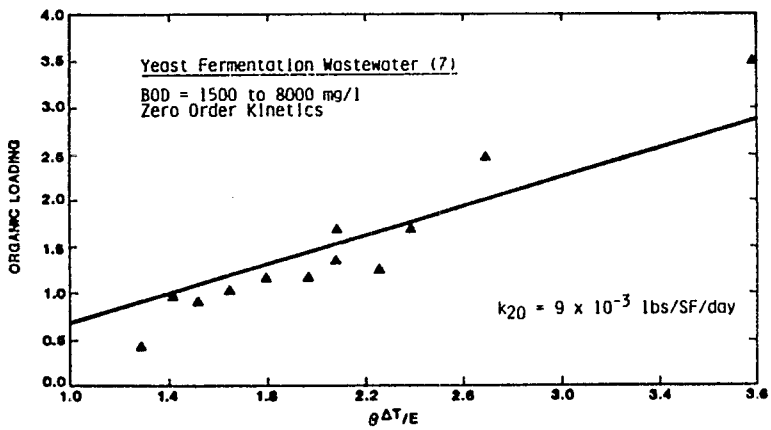

Figure 8 Yeast Fermentation Wastewater Treated on an Incline Plane (7)

3. Surface area characteristics of packing media.

4. Hydraulic characteristics of packing media.

Hydraulic loading on a full scale tower is related to plane hydraulics by geometry as follows:

$$q = (U) \quad (A''_v)$$

where:

q = application rate to tower in gal min^{-1} ft^{-2} of tower area

U = application rate to plane in gal min^{-1} ft of film width

A''_v = slimed area of tower packing media supporting a film growth - $ft^2 ft^{-3}$

Slime thickness reduces exposed surface area below that available from bare media. A knowledge of media configuration and slime thickness can be employed to determine the correction required as follows:

$$A''_v = (A'_v) \ (f_t)$$

where:

A'_v = wetted area of tower packing media supporting a film growth - $ft^2 ft^{-3}$

f_t = a factor for the reduction of slime area below that of bare media due to thickness of slime growth

Laboratory observations of film thickness indicate an f_t range of 0.80 to 0.90. Hydraulic characteristics of packing media are introduced by relating the wetted surface area A'_v and detention time to hydraulic loading. Adjustments of this type are required primarily for media shape and an "aspacked" geometry other than that obtained with a vertically oriented media. The adjustments account for the change in wetted area which occurs as liquid impinges upon randomly packed media and is splashed or otherwise diverted into contact with additional media surface which would otherwise remain unslimed. Adjustment can also be made for the change in reaction rate as a result of a change in the rate of transport of BOD from the flowing liquid to the slime surface. This can also include the effects of removal of suspended BOD by agglomeration

processes. These additions considerably complicate the mathematical descriptions of the process while many times failing to increase the accuracy of design calculations.

For randomly packed media, adjustments for hydraulic effects can be made using a mathematical form prevalent in chemical engineering when packed towers are analyzed, i.e.

$$\frac{A'_v}{A_v} \backsim (q)^n$$

and

$$\frac{(k')}{(k)} \backsim (q)^n$$

Substitution of the above scale-up relationships into the equation for plane performance under first order kinetics yields the equation for full scale performance as follows:

$$f = e^{CK'_{20}\Theta^{\Delta T}} A_v f_t {}^{H}\!/_{(q)^n} \quad (19)$$

where:

C = a combined constant for all hydraulic effects.

The value of the exponent (n) will vary from an expected minimum of 0.50 for randomly packed media to 1.0 for packing similar to stacked vertical sheets. For vertical media a sheet flow regimen dynamically similar to that of plane hydraulics tends to be maintained. However, an adjustment is required for use of less than total media area resulting from distribution hydraulics through the tower. A constant adjustment of 90 percent media utilization is employed as follows:

$$\frac{A'_v}{A_v} = C_w = .90$$

where:

A'_v = wetted surface area ($ft^2 ft^{-3}$)

A_v = manufacturer's rating for dry media ($ft^2 ft^{-3}$)

C_w = a coefficient for wetting efficiency

for vertically oriented media the value of the hydraulic coefficient (C) in equation (19) equals C_w and the design relationship is stated as follows:

$$f = e^{K_{20}\theta^{\Delta T_H}/(q)^n} \qquad (20)$$

where:

$$k_{20} = k'_{20} \cdot A_v \cdot f_t \cdot C_w$$

The above scale-up procedure assumes that the adjustment for non-uniform slime growth can be made by using a single linear correction factor (C_w). This is equivalent to assuming that areas of non-slime growth occur with equal size and frequency throughout the tower depth. This assumption may or may not apply to all tower designs. The alternative approach is to assume that the uniformity of hydraulic distribution will vary in a non-linear fashion with tower height (H). In this circumstance the effective wetted surface area will be related to height as follows:

$$\frac{A'v}{Av} = C\ (H)^{1-m} \qquad (21)$$

where:

C = a correlation constant

(1-m) = a measure of the non-linear distribution of film area.

As the value of m increases, the non-uniformity of film growth with height also increases. At m = 0 slime growth is uniform throughout the tower at m = 1 there is no slime growth in the tower. The full scale relationships when m > 0 is:

$$f = e^{k_{20}\theta^{\Delta T}(H)^{1-m}/(q)^n} \qquad (22)$$

Full scale design equations for the balance of the kinetic models are summarized on Table 1.

Scale-up Calculations

Whey and Sewage - using a packing with $A_v = 27\ ft^2ft^{-3}$

k'_{20^oC} plane = 0.0018 gpm/ft^2 = k'_p

$A'_v = Cw \times f_t \times A_v$

$= 0.9 \times 0.9 \times 27 = 22\ ft^2ft^{-3}$

$k_{20^o(TOWER)} = k'_p\ A'_v$

$$= 0.0018 \text{ gpm/}_{ft^2} \cdot 22 \frac{ft^2}{ft^3}$$

$$= 0.04 \text{ gpm/ft}^3$$

Observed $k_{20^o} = 0.03$ gpm/ft^3

Yeast Wastewater - using a packing with $A_v = 27$ ft^2 ft^{-3}

k'_{20^o} plane $= 0.009$ lbs/day/ft^2

$A_v = 27$ ft^2/ft^3

$A''_v = 22$ ft^2/ft^3

$k_{20^o \text{ TOWER}} = 0.009 \times 22 = 0.198$ lbs/day/ft^3

Observed $k_{20^o} = 0.180$ lbs/day/ft^3

Table 1

Performance Relationships for Packed Tower Trickling Filters

Reaction Model	Relationship @ n = 1.0	Relationship @ n<1 & m>o	
Zero Order	$1/f = 1 - k/(F/M)$	$1/f = 1 - \frac{k(H)^{1-m}}{(S_a)(q)^n}$	(23)
First Order	$f = e^{k\,H/q}$	$f = e^{k(H)^{1-m}/(q)^n}$	(24)
Simple Retardant	$f = 1 + k\,H/q$	$f = 1 + k(H)^{1-m}/(q)^n$	(25)
Exponential Retardant	$f = 1 + \frac{NkH}{q}$	$[f] = 1 + \frac{Nk(H)^{1-m}}{(q)^n}$	(26)
Concentration Dependent	$f = e^{k/(F/M)}$	$f = e^{\frac{k(H)^{1-m}}{(Sa)(q)^n}}$	(27)

Full Scale Performance

For a given type of packing, tower volume can vary with the following design parameters:

1. Liquid application rate
2. Recycle ratio
3. Tower height
4. Efficiency of BOD removal

The effects of variations in the first three parameters are dependent on (1) the necessity to maintain a minimum wetting rate and (2) the numerical value of the exponent (n). In all cases, an increase in efficiency of removal requires an increase in tower volume. In general, the effects of design parameters can be described as follows:

Design Variables	Change in Variable	Change in tower volume
H	Increase	Decrease or no ch
r	Increase	Increase or no ch
q	Increase	Increase or no ch
E	Increase	Increase

Structural requirements and hydraulic distribution problems limit maximum tower heights. Heights of 20 ft are common with maximums to 45 ft.

Commercial packing of the lattice structure type appears to require minimum application velocities of $\pm$ 0.5 min ft^{-2}. Operation below the minimum velocity can result in progressively less utilization of tower packing.

In order to maintain commonly used heights and provide a minimum application velocity, effluent recycle is usually required. The added tower volume required to accommodate the recycle varies with both the efficiency of removal sought and the recycle ratio finally employed. Because of the

non-uniform influence of process variables, design calculations can involve relatively complex manipulations.

Process design data from pilot plant operation are obtained using relationships (23) to (27) together with graphical correlation techniques. The type of correlations used while mathematically identical can non-the-less effect the value of the process constants obtained and, therefore, the design of the full scale unit. Techniques which utilize separate rearrangements of Equations (23) to (27) to calculate m & n independently can have this tendency. A single correlation equation can be obtained by taking logarithum's twice and rearranging the result to provide a linear relationship. When m = o the full scale first order equation is expressed as:

$$\log\left[\frac{H\theta^{\Delta T}}{2.3\ \log\ (f)}\right] = n\ \log(q) + \log\ 1/k_{20} \quad (28)$$

The slope of the correlation is (n) and the intercept at q = 1.0 is log $(1/k_{20})$.

In order to provide for the possibility that m > 0 a two part technique is used and is illustrated Figure 9.

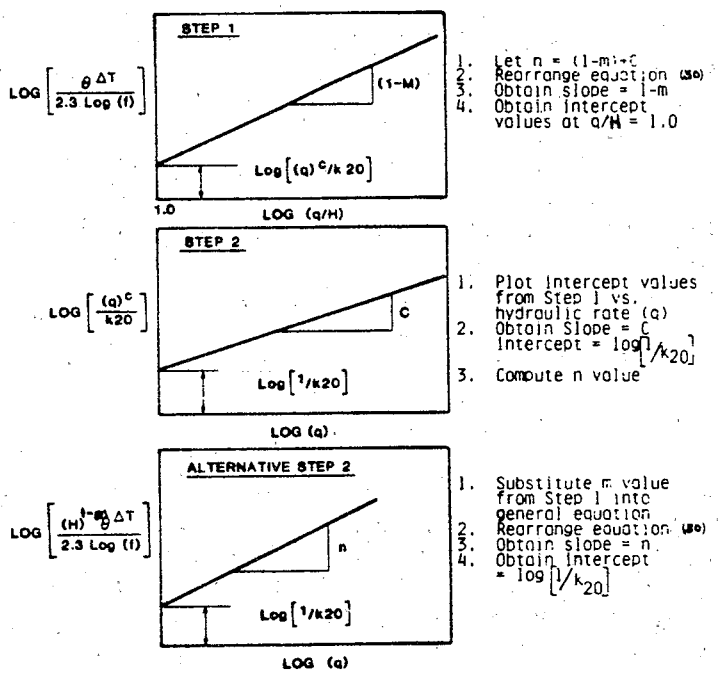

Figure 9 Graphical Analysis for m greater than o

A correlation is first obtained for variable tower height at constant hydraulic rate (q). Double logarithums of equation (24) are used together with a mathematical substitution designed to eliminate the effects of the hydraulic coefficient (n) from the correlation. With the influence of (n) thus eliminated, the value of m may be determined. The elimination of (n) is obtained by making the following substitution,

$$n = (1-m) + C \qquad (29)$$

where C = a constant whose value, determined from test data, will be such as to make the above statement valid.
Using this substitution with equation (24) the following linear equation is obtained after logarithums are taken twice:

$$\log \left[\frac{\Theta^{\Delta T}}{2.3 \log (f)}\right] = (1-m) \log (q/H) + \log (\text{CONSTANT}) \qquad (30)$$

where:

$$\text{Constant} = (q)^c/k_{20}$$

Using variable height data with constant hydraulic rate a plot of equation (30) will yield (1-m) as a slope. At this juncture, no use of the correlation intercept is made as the purpose of the substitution was to remove (n) from the slope and relocate it in a constant value intercept as part of C. The intercept value has no other use in this part of the correlation.

If data are taken over a range of hydraulic rates then a plot may be made separately for each (q) value. This will provide a number of intercept values. These intercept values may then be related to hydraulic rate in such a way as to yield the numerical value of C.

$$\log \left[\frac{(q)^c}{k_{20}}\right] = C \log(q) + \log 1/k_{20} \qquad (31)$$

Log $(q)^c/k_{20}$ is obtained as the intercept value from prior use of equation (30). Tower performance on the treatment of black liquor from Eckenfelder (8) is shown in Figure 10. Data for an integrated kraft mill effluent, Quirk (9), is shown on Figure 11. Again n = 1.0 and a k_{20} = 0.056 gpm/CF is determined.

Experience of the authors has shown that unless tower operation is below a minimum wetting rate or there is a

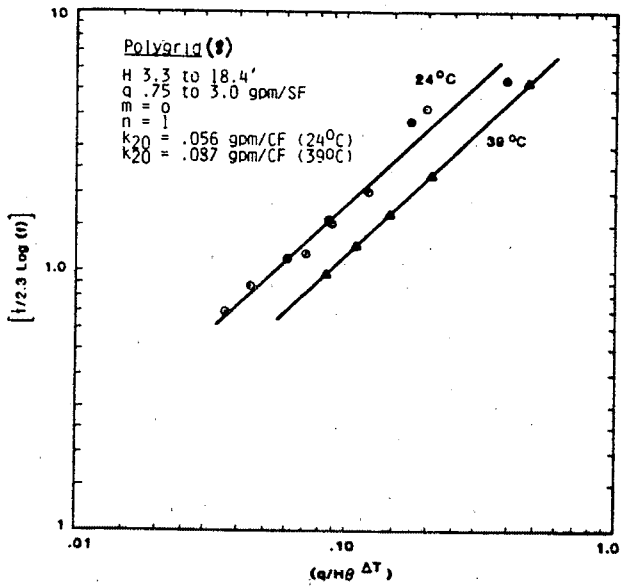

Figure 10 Treatment of Black Liquor on a Polygrid Tower (8)

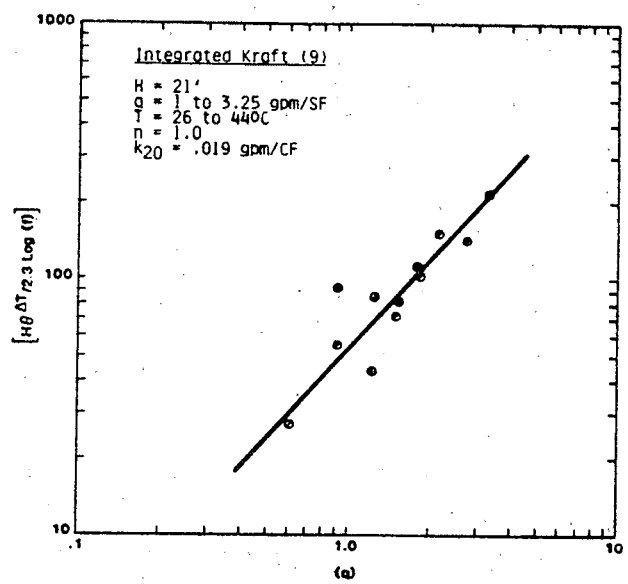

Figure 11 Treatment of Integrated Kraft Effluent on a Surfpac Tower (9)

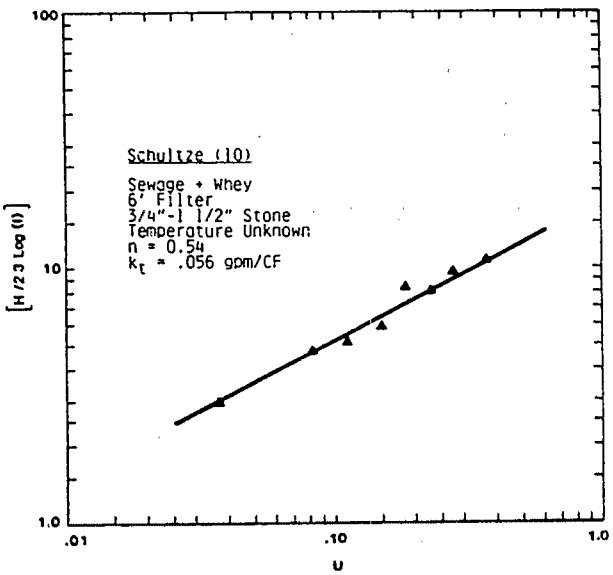

Figure 12 Whey and Sewage Treated on a Gravel Packed Filter (10)

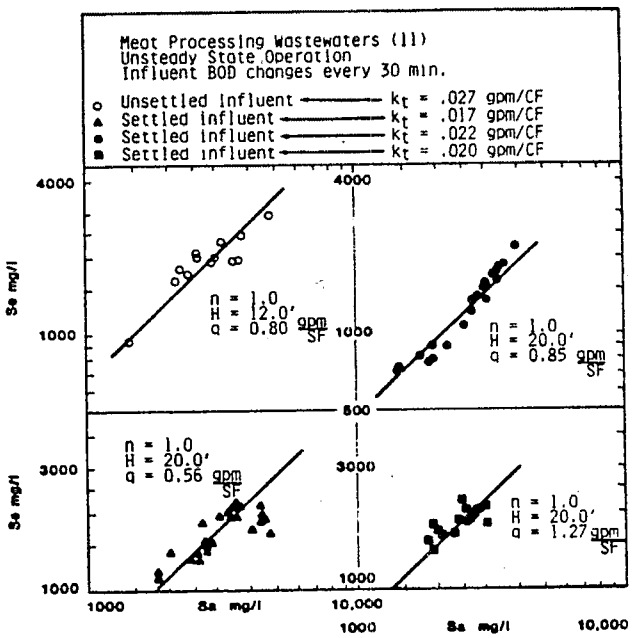

Figure 13 Unsteady State Treatment of Meat Processing Wastewaters on a 20 ft. Tower (11)

malfunction in the hydraulic distribution, the value of m will equal 0 and n = 1.0 when dealing with vertical sheet flow packing.

When working with random packing, values of $n < 1.0$ will be obtained. The data of Schultze (10) treating sewage plus whey on a 6'deep gravel packed filter are shown on Figure 12. The 3/4 to 1 1/2" stone media provided a specific surface area computed at 36 SF/CF. Filter operation covered a range in hydraulic loading of from 0.0375 to .375 gpm/SF. An (n) value less than 1.0 is evident. A reaction rate of .056 gpm/CF was determined.

The doctoral thesis data of Moodie (11) allows an examination of the correlation technique when applied to unsteady state performance. Using a 3 sq/ft tower, 20' high, Moodie experimented with meat processing wastewaters with a COD of up to 5000 mg/l. Hydraulic rates varied from 0.56 to 1.27 gpm/SF and an intermediate effluent sampling point at 12' was used. Every 30 minutes the COD of the influent was changed while the hydraulic rate remained constant. The response of the tower to these variations in loading is shown in Figure 13. Effluent concentration (S_e) is related to influent concentration (S_a) using a First Order Reaction and equation (24). Even though there is considerable scatter in some of the data, the general applicability of the correlation is evident.

A comparison of unsteady state reaction rates with hydraulic loading and influent concentration variations is proivded in Table 2 and demonstrates that with careful sampling technique, stable performance data can be obtained from unsteady-state operation. Where such control can not be obtained, as in a field installation which is not used for a Ph.D. thesis, much more variation in unsteady-state performance can be expected. This is illustrated in Figure 14 for yeast fermentation effluent treated in a 21 pilot tower receiving continuously variable flows and influent concentrations, Quirk (7). The inclined plane data previously referenced on Figure 8 were superimposed on the pilot plant data. The scale-up calculations described above were used. The laboratory k' of 9×10^{-3} scaled-up to 0.180 lbs/CF/day. Figure 14 illustrates that on the average the unsteady-state field data group around the scale-up line from the laboratory plane.

An additional aspect of hydraulic loading is illustrated in studies of Eckenfelder (15). Eckenfelder utilized a pilot plant equipped with a vertically oriented asbestos packing, operating over 7 to 1 height change, and 3 to 1 variation in hydraulic loading. His data are presented in Figure 15.

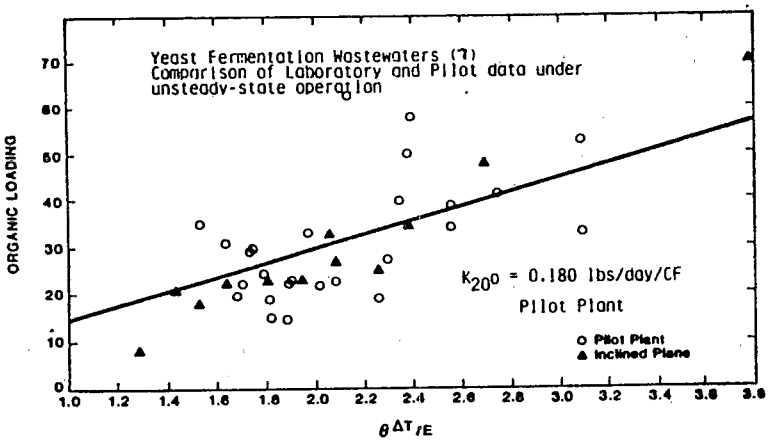

Figure 14 Comparison of treatment of Yeast Fermentation Wastewaters on a 21 ft. Surfpac Tower (7)

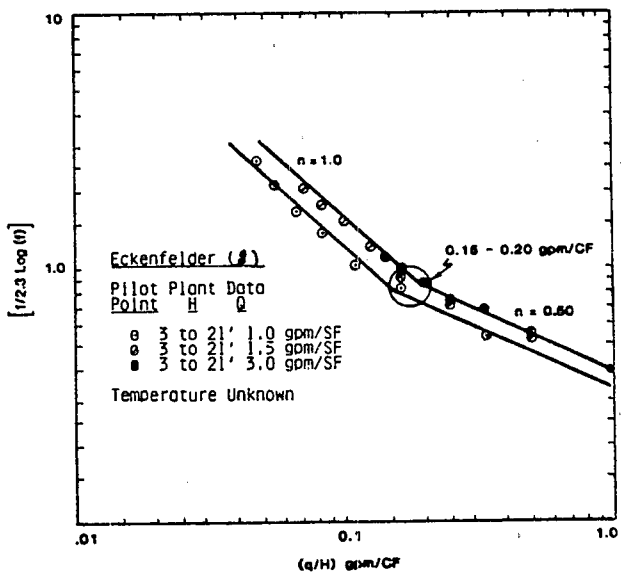

Figure 15 Black Liquor Treatment on an Asbestoes Tower (8)

Table 2

Applications of First Order Reaction Correlation to Unsteady-State Operation after Moodie (11)

Hydraulic Rate	Tower Height	COD Range	Reaction Rate
0.56 gpm/SF	20.0'	1750-4800 mg/1	0.017 gpm/CF
0.85 gpm/SF	20.0'	1400-4000 mg/1	0.022 gpm/CF
1.27 gpm/SF	20.0'	1750-3000 mg/1	0.020 gpm/CF

It is seen that a volumetric loading rate of between 0.15 and 0.20 gpm/CF on abrupt change occurs in the value of the hydraulic coefficient (n). Above this threshold value and (n) of 1.0 fit the data quite well. Below it the n decreased to .50. This indicates that at some limiting value of hydraulic loading n will decrease below 1.0 in response to poor wetting efficiency and flow channeling which prevents uniform slime growth throughout the tower.

Process Design

The previous illustrations of correlation approaches and data fitting procedures underscore the fact that determination of the proper model is of paramount importance. In a similar manner, it is necessary that laboratory and/or pilot scale studies incorporate the full range of design variables. Extraplotation beyond the confines of measured data can be risky at best.

The effect of recirculation on design capability is a topic of particular concern in process design. Conflicting data and opinions populate the literature. Recirculation decreases the detention time while increasing the velocity of flow &turbulent transport of BOD to the slime/liquid interface which are opposing effects. A minimum hydraulic loading is necessary for through wetting and this can be provided by recycle. In order to approach the recycle question properly it is necessary that recirculation over a wide range be utilized in treatability studies so that data exist for recycle ratios in excess of those ultimately selected for final design. With these data in hand, process models may be employed to determine

the net effect of recirculation.

Using the first order model as an illustration, the process design calculation re-express the recycle/efficiency function (f) as follows:

$$f = \frac{1 + r\ (1-E)}{(1+r)\ (1-E)} = \frac{Ec + r}{1+r} \qquad (32)$$

Where E_c relates to BOD removal efficiency i.e.,

$$E_c = (1/1-E)$$

and allows topography to be simplified as the design analysis procedes. The concept of a unit volume of tower per unit of raw wastewater flow is then introduced. Equation (33) is employed to obtain a ratio of these unit volumes with and without recirculation. The final relationship between unit volumes is shown below:

$$\frac{V}{V_o} = (1 + r)\ (C_{E,r}) \qquad (33)$$

where:

V = the unit volume when recirculation is practiced

V_o = the unit volume without recirculation

$C_{E,r}$ = a sensitivity constant for a given efficiency and recycle ratio

When the constant $(C_{E,r})$ equals its maximum value of 1.0, the unit volume required with recirculation varies directly in proportion with the arithmetic effect of recycle i.e. $\frac{V}{V_o} = (1 + r)$. At constant values less than 1.0 the increase in tower volume is reduced below that dictated by hydraulic through-put ratio. This metigation effect is formulated for a first order reaction as follows:

$$C_{E,r} = \left[\frac{\log \frac{(E_c+r)}{1+r}}{\log (E_c)}\right]^{1/n} \qquad (34)$$

Using a vertically oriented media and an n value of 1.0, the value of $(C_{E,r})$ will vary with design efficiency as illustrated on Figure 16. While the impact of recirculation is reduced on Figure 16 as recycle ratio increases, total tower volume

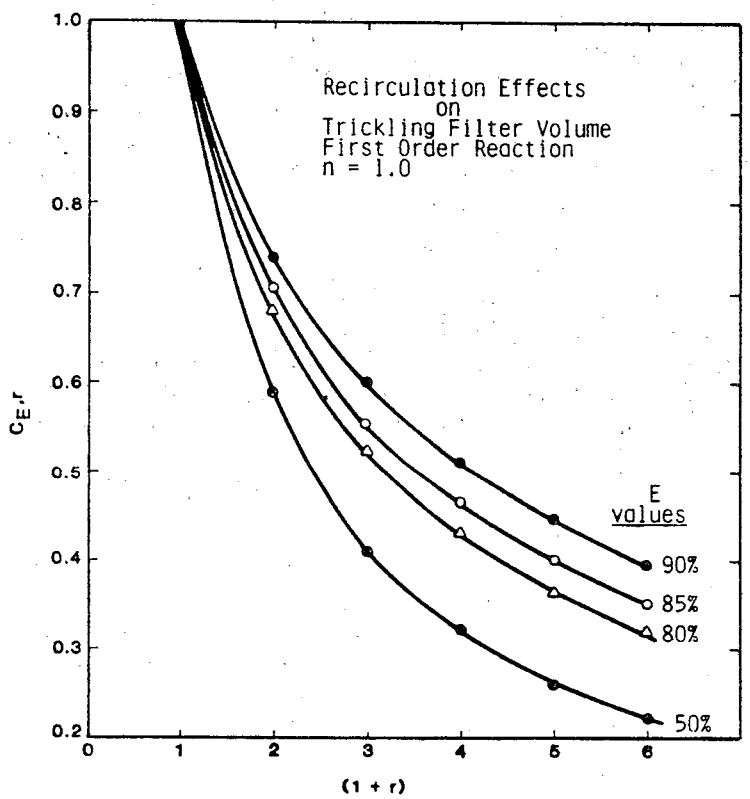

Figure 16 Recirculation Effects on Trickling Filter Volumes

experiences a net increase. Unit volume ratios are compared with design efficiency requirements and recycle ratios in a general manner on Table 3. The analysis shows that recirculation effects are moderate only for low removal efficiency or tower performance as a pretreatment or roughing unit.

At (n) values less than 1.0 the effects of recycle are much reduced over these shown on Table 5. For example, at n = 0.50 the unit volume ratio approaches 1.0 even though efficiency is 90% and a recycle ratio of 2.0 is used. This may be compared with a volume ratio of 1.7 under comparable conditions when n = 1.0. However, this can be only an apparant volume reduction in that at n values less than 1.0 tower media is not being used effectively and the value of k_{20} will be reduced. This can offset the reduced effects of recycle usage.

A similar process design approach is used for the alternative reaction models. Unit volume relationships for these models are shown on Table 4.

A summary of reaction rates and reaction models for industrial effluents is presented on Table 5. In each case sited, tower media was of the vertically oriented type.

Table 3

Effect of Recirculation

on

Tower Unit Volume

First Order Reaction

n = 1.0

Efficiency %	Recycle Ratio(r)	Unit Volume Ratio (v/v_o)
90	1	1.48
90	2	1.80
90	3	2.04
90	4	2.25
50	1	1.18
50	2	1.23
50	3	1.29
50	4	1.30

Table 4

Effect of Recirculation

on

Tower Unit Volumes

Reaction Models	Basic Multiplier $n = 1$	Basic Multiplier $n = n$	Efficiency Recycle Constant $n = 1$	Recycle Constant $n = n$
1. Zero Order	1.0	1.0	1.0	1.0
2. First Order	(1+r)	(1+r)	$\left[\frac{\log(\frac{E_c+r}{1+r})}{\log\ (E_c)}\right]$	$\left[\frac{\log(\frac{E_c+r}{1+r})}{\log\ (E_c)}\right]^{1/n}$
3. Simple Retardant	1.0	1.0	1.0	1.0
4. Concentration Dependent	(1+r)	(1+r)	$\left[\frac{\log\left(\frac{E_c+r}{1+r}\right)}{\log\ (E_c)}\right]^{1/n}$	$\left[\frac{E_c+r}{E_c(1+r)}\right]^{1/n}$

Table 5

Comparison of Rate Constants Trickling Filtration of Industrial Effluents

Effluent	Reactor	Model	n	k_{20}
1. Integrated Kraft Mill				
mill A	Tower	First Order	1.0	0.021 gpm/CF
" B	"	"	"	0.022 "
" C	"	"	"	0.017 "
" D	"	"	"	0.014 "
" E	"	"	"	0.018 "
" F	"	"	"	0.034 "
" G	"	"	"	0.044 "
2. Rag Mill Effluent	Tower	First Order	1.0	0.083 "
3. Box board Mill Waste	Tower	First Order	1.0	0.027 "
4. Yeast Fermentation Wastewater	Tower	Zero Order	1.0	.180 lbs/day/CF
5. Yeast Fermentation Wastewater	Plane	Zero Order	1.0	.009 lbs/day/CF

6. Meat Processing effluents	Tower	First Order	1.0	.020 gpm/CF
7. Dilute Black Liquor	Tower	First Order	1.0	.056 gpm/CF
8. Canning Wastes	Tower	First Order	1.0	0.021 gpm/CF
9. Slaughter House	Tower	First Order	1.0	0.044 gpm/CF
10. Whey plus Sewage	Stone	First Order	1.0	0.056 gpm/CF
11. Whey plus Sewage	Plane	First Order	1.0	.0018 gpm/SF
12. Whey plus Sewage	Screen	First Order	1.0	.0025 gpm/SF
13. Pharmaceutical effluents	Plane	Concentration Dependent	1.0	2.9 lbs/SF/day
14. Insulating Board Waste	Plane	Zero Order	1.0	.005 lbs/SF/day

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TREATMENT OF COKE PLANT WASTEWATERS IN PACKED BED REACTORS

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The treatment of wastewaters originating from coke oven operations has been studied in a wide variety of biological treatment operations. Most of these studies have shown that this wastewater is biodegradable to a large extent and that given enough acclimation, the bacteria can develop to fully treat compounds like phenol, cyanides and thiocyanates. The work that will be presented in this paper describes the results of a treatability study of wastewaters originating from a benzol plant in an upflow biotower (UBT). The wastewater constituents are the same as for coke oven wastewater, except that most of the constituents are present in somewhat lower concentrations. The biotower used in this study is a biological treatment process developed by the Leopold Company and operates more or less like a reversed flow trickling filter. The tower is packed with random plastic medium, the influent flows upward and air is supplied by aeration through a filter underdrain system. This paper will present the results of the treatability of benzol plant wastewater in this reactor. The main purpose of this study was to determine the loading at which the phenol was virtually completely removed from the influent. The data obtained in this study will be

compared with studies performed by other researchers with similar wastewater in activated sludge and with other types of fixed film reactors. In this way, it will be possible to compare the performance of the different biological treatment reactors available.

EXPERIMENTAL SET-UP

The work conducted in this project was done with a pilot unit of the upflow biotower (UBT). Figure 1 is a schematic diagram of the pilot plant set-up. The tower had dimensions of 2 x 3 x 10 ft. high and was filled with 39 $ft.^3$ of random plastic medium with a specific surface area of 30 ft^2/ft^3. The air was distributed through a filter underdrain at a rate of 2-5 $scfm/ft^2$. The wastewater was fed at varying flow rates (0.5-2 gpm) through the tower. The waste characteristics of the influent to the UBT are shown in Table I. It shows that the average soluble organic carbon (SOC) concentration is 435 mg/l. The COD is about 1500 mg/l. This strength is about half of waste ammonia liquor. The average phenol concentration was only 36 mg/l, which is significantly less than in waste ammonia liquor. The fact that the ratio between COD and phenol in this waste is so much higher than in normal waste ammonium liquor, indicates that the waste from the benzol plant contains many more organics in addition to phenol, while waste ammonia liquor primarily contains phenols as the organic material. Other specific organics analyzed to be present in this waste are benzene, toluene and napthalene. The concentrations from grab samples for these specific organics during this study are also shown in Table I. The pilot plant was started up by feeding activated sludge from an existing wastewater treatment plant treating coke oven wastewater. The phenol removal was virtually complete a few days after adding seed to the tower. During the pilot study, the reactor was fed hydraulically with 0.5, 1 and 2 gpm of wastewater. The influent values fluctuated throughout this period so that the organic load also fluctuated from day to day.

RESULTS OF THIS STUDY

The results of this study showed that the organics present in this wastewater are very well biodegradable and that it is possible to treat this wastewater biologically

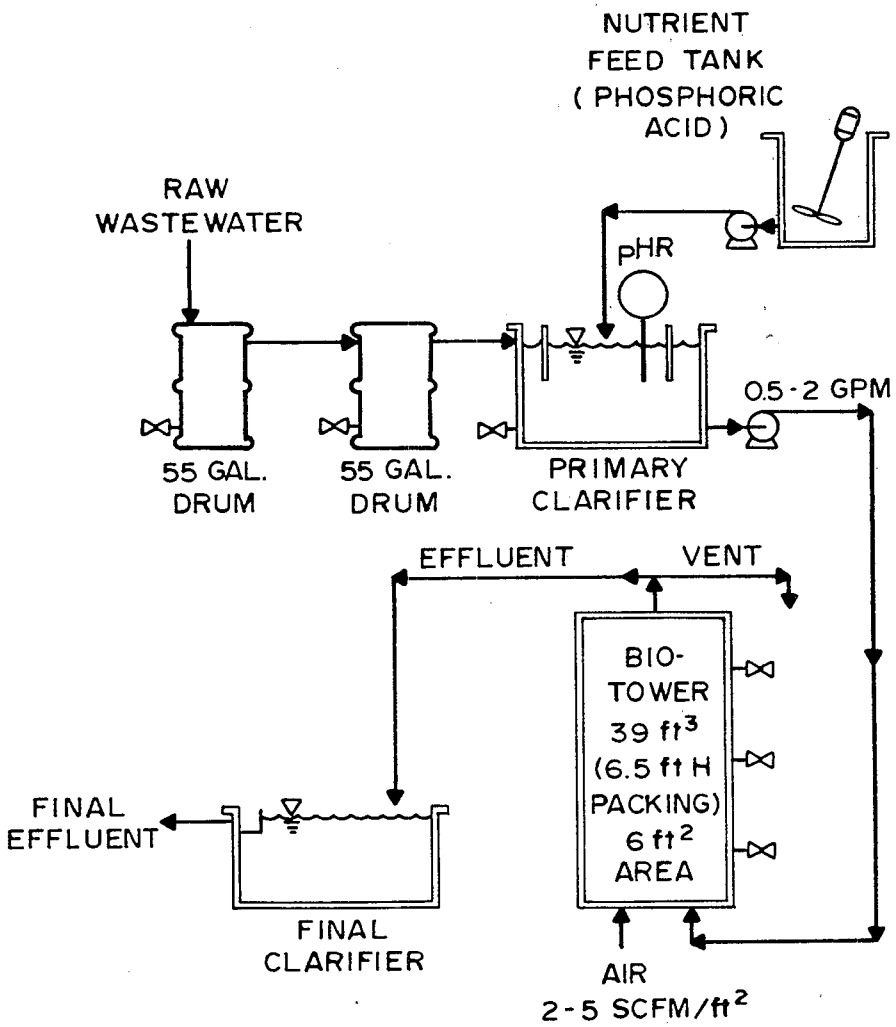

FIGURE I

LAYOUT OF THE PILOT PLANT

TABLE I

INFLUENT CHARACTERISTICS OF WASTEWATER TO UBT

Parameter	50 % Value*	90% Value*
Suspended Solids	50	N/A
Oil and Grease	32	200
Phenol	36	67
SOC (Soluble Organic Carbon)	435	630
SCN	63	102
CN - Free	11.3	18
CN - Total	21.8	33.4
BOD_5	700	995
COD	1500	2300
NH_4-N	42	85
Sulfides	40	64
Grab Sample		
Benzene	208	
Toluene	30	
Naphtalene	8	

* 50% Value: Mean concentration from a probability plot.

90% Value: Indicates the concentration that is not exceeded in 90% of the samples analyzed.

to a very high extent. The probability curve for the percentage SOC removal is shown in Figure 2. Typical removal percentages for the various parameters are shown in Table II. The hydraulic load was 1 gpm/39 ft^3 or 37 gpd/ft^3 reactor. The corresponding average organic load was 120 lb SOC/1000 ft^3 day or 400 lb COD/ 1000 $ft^3 \cdot day$.

TABLE II

PILOT BIOTOWER LOADINGS AND OBTAINED REMOVALS DURING RUN III - 1 GPM

Parameter	Mean Loading $lbs/1000\ ft^3 \cdot d$	Mean Removal (%)
SOC	135	57
COD_f	462	63
$BOD_{5,f}$	216	51
CN-Free-Filtered	3.5	71
CN-Total-Filtered	6.7	75.2
Phenols-Filtered	11.1	99.9±
SCN-Filtered	19.5	61
NH_4^+	13	0
Sulfides	4.2	73

In Figure 3 the percentage SOC removal is plotted versus the effluent phenol concentration. This shows that as long as at least 40-50% of the SOC is removed, the effluent phenol concentration is very low. The removal of free cyanides was not complete at the loadings at which the plant was operated. Based on performance of other systems, however, it is felt that free cyanide is totally

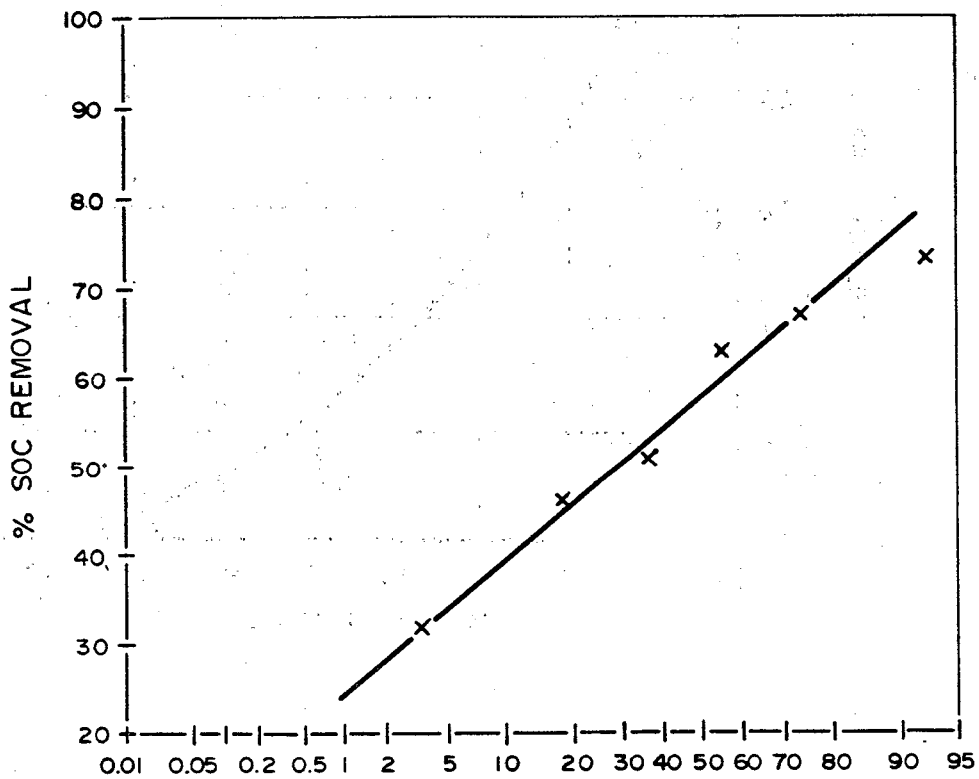

FIGURE 2

CUMULATIVE FREQUENCY DISTRIBUTION
OF PERCENTAGE SOC REMOVAL
DURING RUN III

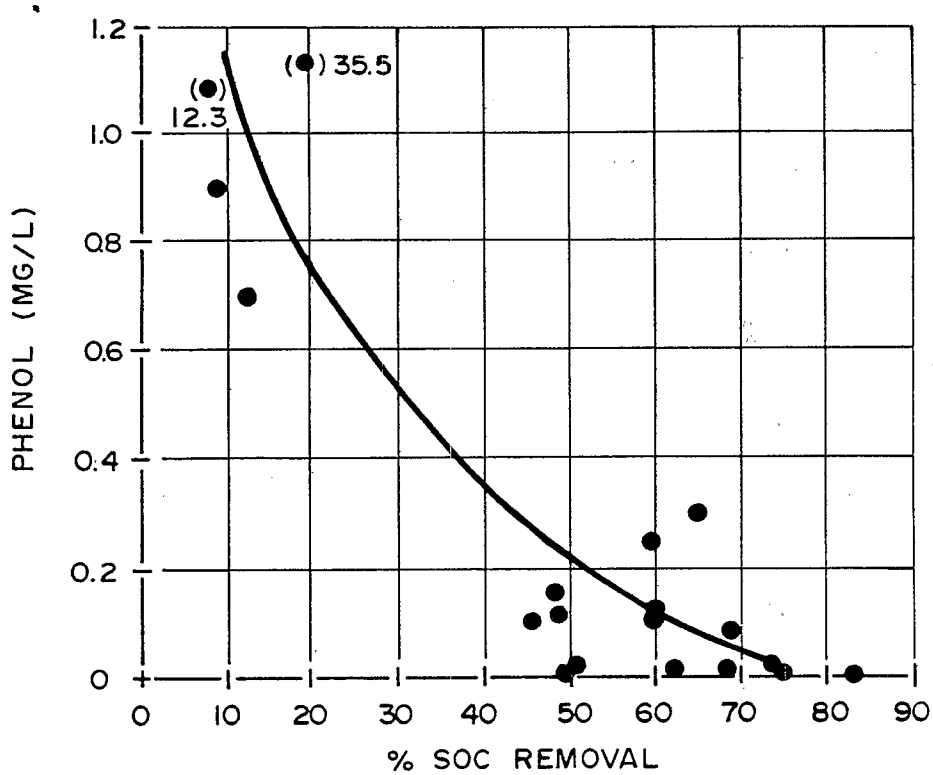

FIGURE 3

EFFLUENT PHENOL CONCENTRATION
AS A FUNCTION OF
PERCENTAGE SOC REMOVAL

biodegradable, provided the loading is low enough. The same is true for thiocyanates. (See Discussion section.)

The only nutrient required for this waste stream was phosphorus. This was added at such a concentration that the effluent phosphorus concentration was maintained at 1 - 2 mg/l. This required on the average only 5 mg/l.

The amount of sludge generated in this system represented a yield factor based on a BOD_5 basis of about 50%. This is comparable to conventional high-rate biological systems. The loading at which this sludge yield was obtained was quite high (216 lb BOD_5/1000 ft^3/day).

The air requirements for this unit were comparable to activated sludge. The DO concentration in the effluent was normally high (6-7 mg/l) and therefore in the scale-up of the system, it is possible to reduce the amount of air required under normal operating conditions.

Throughout the course of this six-month study, it never proved to be necessary to backflush the tower. The excess sludge sloughed off at a sufficient rate to maintain a good flow and aeration throughout the system. The upflow flowrate through the tower was about 0.16 gpm/ft^2.

Throughout this project, the unit suffered several shock loads as result of leaks or spills in the plant from where the waste originated. The unit showed remarkable potential for quick recovery. Figure 4 shows a few cases where the SOC jumped up from one day to another and the effluent SOC value also jumped up. However, after stabilizing the influent condition, the effluent of the biotower recovered rapidly. Throughout this study, the phenol concentration in the effluent was only in two effluent samples above 1 mg/l. In each case, the phenol concentration was below 1 mg/l the following day.

The kinetics of this system can be expressed by the equation

$$S_e/S_o = EXP(-K/L)$$

where S_e is the effluent concentration, S_o is the influent concentration, K is a constant, L is loading. By plotting log S_e/S_o versus 1/L, it is possible to obtain the value of the removal coefficient. This is done in Figure 5 and it shows a removal coefficient of K = 0.18 lb/cu.ft/day.

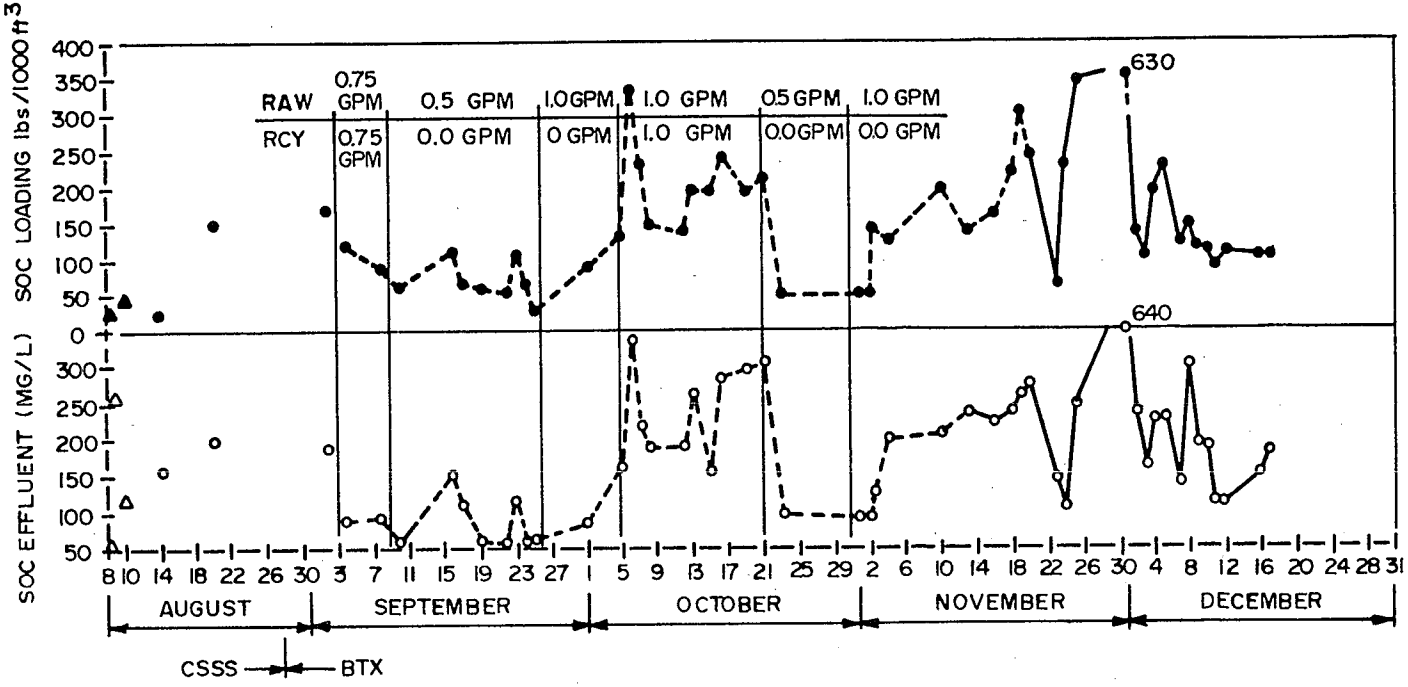

FIGURE 4

HISTOGRAM OF SOC REMOVAL VERSUS LOADING

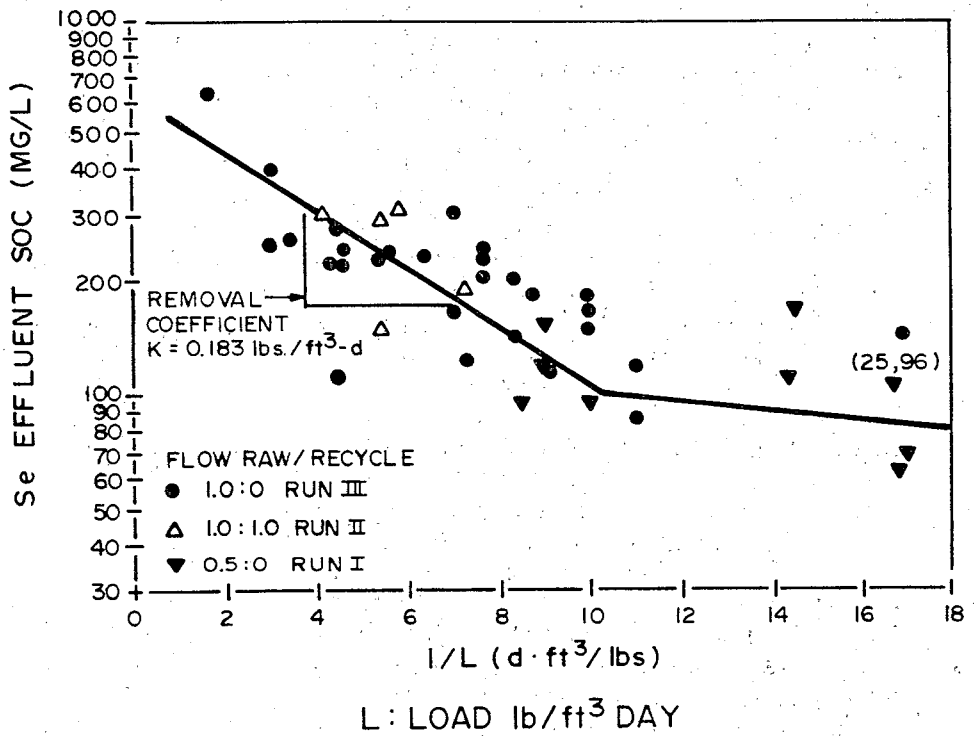

FIGURE 5

KINETICS OF SOC REMOVAL
IN THE PACKED BED REACTOR

DISCUSSION

It is possible to compare the performance of this upflow biotower with results reported by other investigators of other biological systems. Since the main objective of various biological systems is to have the highest organic loading and the smallest possible volume and still provide a stable operation, the results of all data that will be discussed in this paper are expressed in terms of performance versus volumetric organic load. Activated sludge data can be converted to this by assuming that the activated sludge concentration is 3 g/l of active biomass. The results for COD removal are plotted in Figure 6 and it shows that the performance of the UBT is quite favorable. The performance of activated sludge is very close to the packed bed reactor, but one of the disadvantages of activated sludge, i.e. the sludge settling, has been avoided. Also, for an industrial type of operation, it is felt that a fixed film bioreactor is more suitable to handle shock loads.

The removal of CN and SCN was only partial at the organic loadings the plant was operated. Generally, these components are biodegradable at lower loadings. Figures 7 and 8 show the percentages removal and the effluent concentrations for these parameters as a function of loadings for several types of systems. These data show quite a fluctuation between the various reported data. In general, the breakdown of CN and SCN are thought to be totally degradable at lower loadings (<200 lb COD/ 1000 ft^3/day). The CN removal data are more scattered than the SCN data because of inconsistency in reporting CN as total or free. Only the free cyanide is biodegradable and therefore in cases with a large fraction of fixed cyanide, the percentage removal will be low. Why in some cases SCN is not totally removed at low loadings is uncertain. Some research conducted in this area seems to direct the focus to certain environmental conditions (pH, phenol, ammonia concentration) but no total picture is yet available on SCN oxidation (refence 17).

The effluent phenol concentration is of importance since BAT guidelines require very low limits (0.025 mg/l). This is not easily met with a biological system. Figure 9 shows the effluent phenol concentration for various

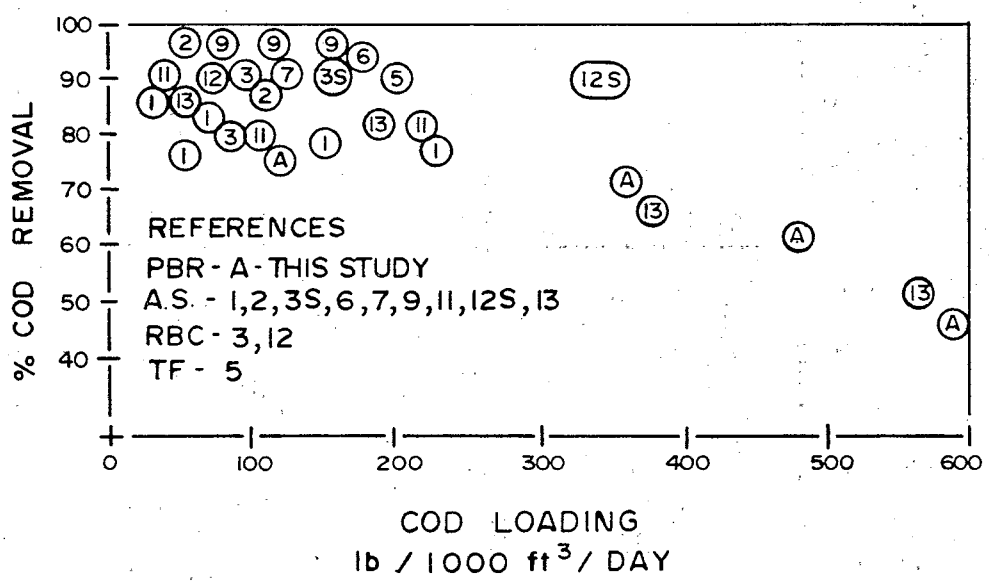

FIGURE 6

COD REMOVAL VERSUS
ORGANIC LOAD FOR VARIOUS
BIOLOGICAL REACTORS

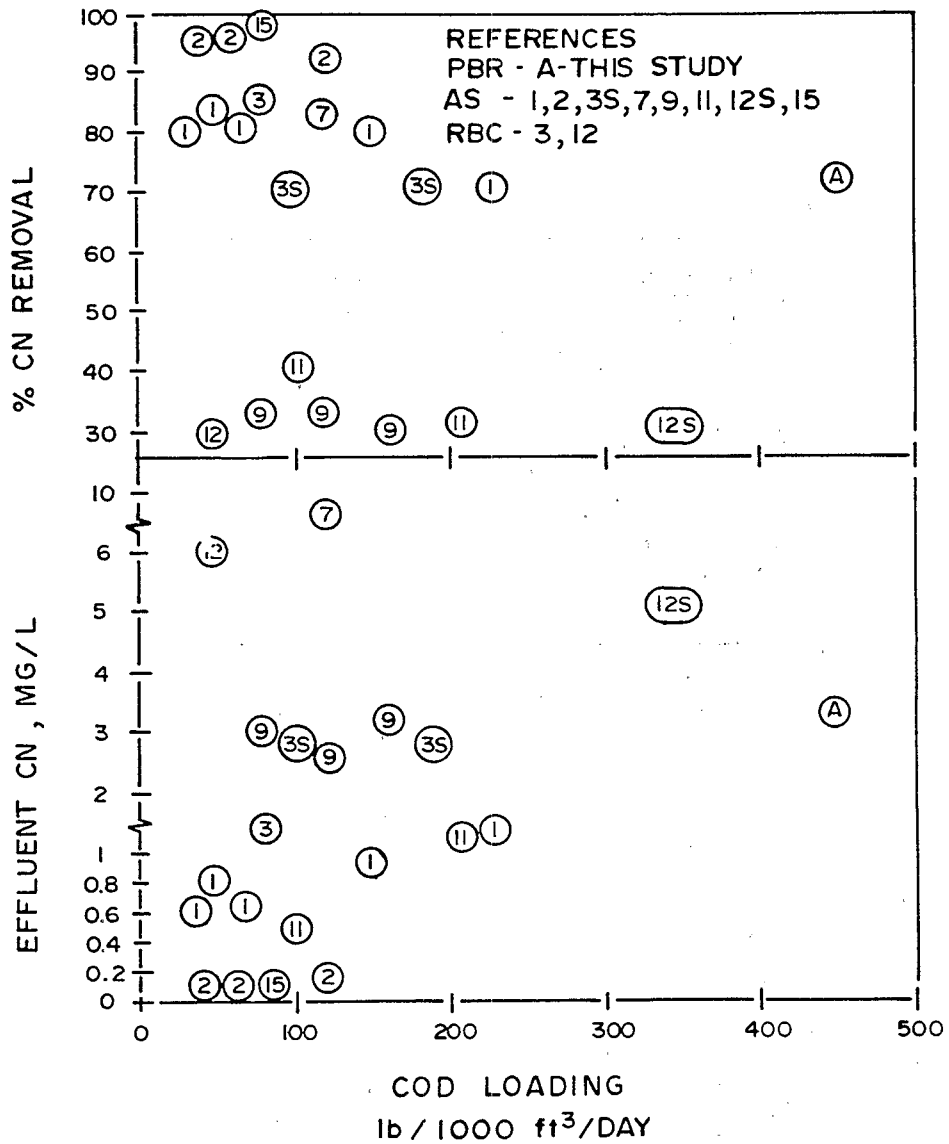

FIGURE 7

CN REMOVAL VERSUS ORGANIC LOAD FOR VARIOUS BIOLOGICAL REACTORS

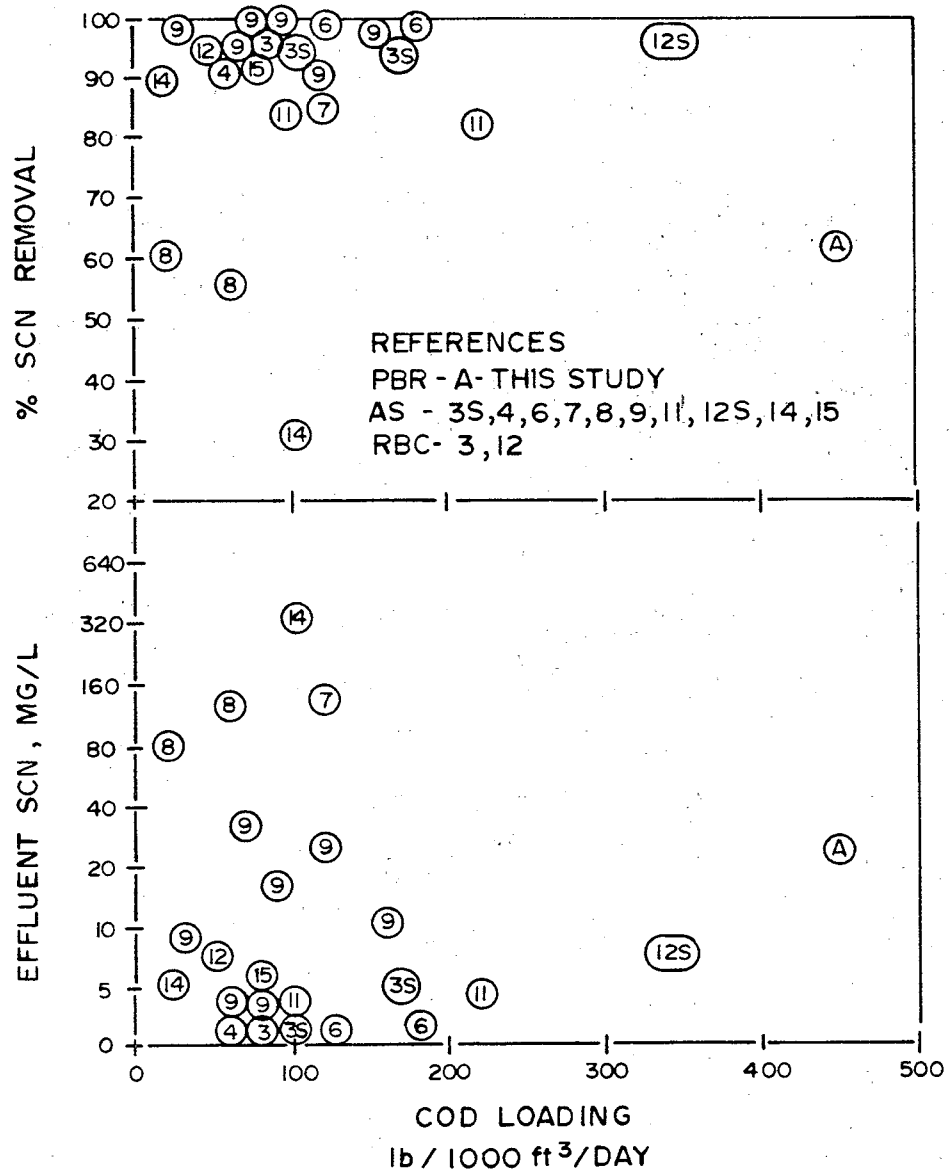

FIGURE 8

SCN REMOVAL VERSUS ORGANIC LOAD FOR VARIOUS BIOLOGICAL REACTORS

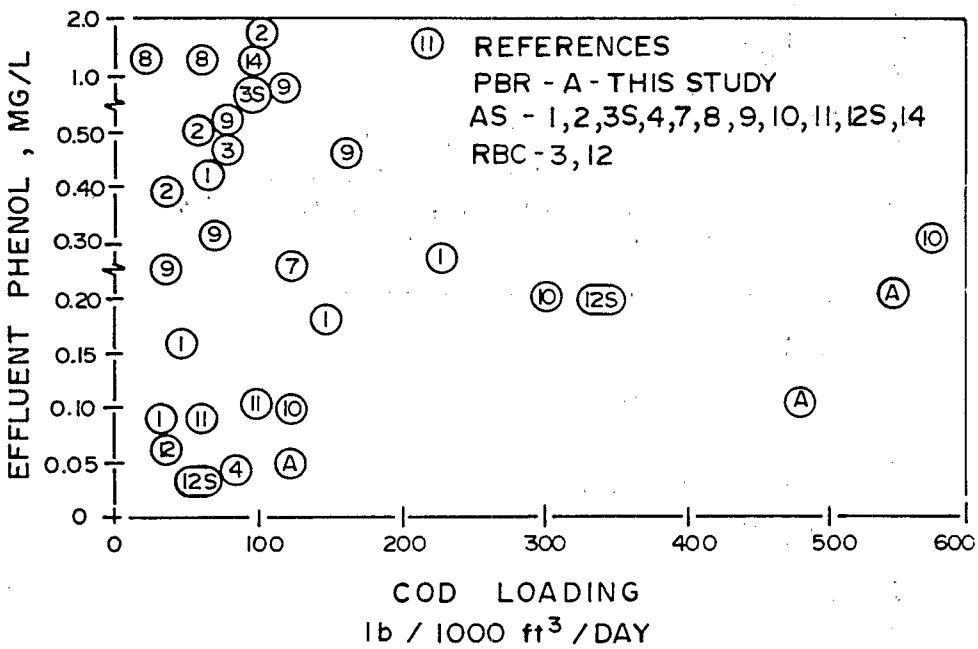

FIGURE 9

PHENOL REMOVAL VERSUS ORGANIC LOAD FOR VARIOUS BIOLOGICAL REACTORS

studies. The figure shows only 3 effluent values below 0.05 mg/l. It is difficult to predict what will be required to get these very low effluent phenol concentrations. Low loadings is obviously a prerequisite but this seems to be no guarantee for low phenol effluent concentration. The analytical techniques used for analyzing phenol could have a large implication at these low residual values.

One factor that requires some additional study is the anticipated improvement in overall performance by staging UBT reactors. As with plug flow versus completely mixed it is felt that by operating 2 reactors in series, the effluent quality will be better than with 1 reactor with the same volume as the 2 reactors combined. With fixed film reactors, there is more of an opportunity for special groups of bacteria to be growing in certain sections of the reactors. In batch tests, it has been shown (reference 16) that the sequence of bio-oxidation of the pollutants in coke oven wastewater is phenol-cyanide-thiocyanate. By staging reactors, it is possible to simulate this type of operation.

Another factor that shows promise for optimizing the performance of the UBT is by reducing the energy requirement per pound of COD removed by increasing the height of the tower. By increasing the height, the air has a longer detention time and the packing should ensure a tortuous path of the air. Both factors seem to work toward an energy efficient reactor. No data are yet available on this but intuitively this approach should be advantageous.

CONCLUSIONS

1. The UBT proved to be an efficient reactor for biological degradation of benzol wastewater. Organic loadings as high as 300 lb COD/1000 ft^3/day resulted in 70% or higher COD removal.

2. The percentage removal of CN and SCN was about 60-70% at the loadings at which the reactor was operated.

3. Phenol is virtually completely oxidized as long as the percentage COD or SOC removal was above 50%. This does not necessarily mean that the phenol is completely oxidized but at least the ring is broken and the resulting organic does not register as phenol.

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TRICKLING FILTER EXPANSION OF POTW BY SNACK FOOD MANUFACTURER

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INTRODUCTION

One of Frito-Lay, Inc.'s largest corn and potato snack food manufacturing plants was recently opened in Killingly, Connecticut, about halfway between Hartford, Connecticut and Providence, Rhode Island. The decision to locate this plant in Killingly required the company to install on-site primary treatment as well as an expansion of the publicly owned treatment work's (POTW's) secondary treatment system. This expansion consisted of constructing a $1.6 million trickling filter and pumping station. This paper describes the joint effort between the town of Killingly and Frito-Lay to design these facilities and obtain State approval to construct them, and reports on system performance to date.

The construction of a trickling filter or a POTW expansion is not unusual. The uniqueness of this plan stems from the fact that this project was the first industrially designed and constructed expansion of a POTW in the State of Connecticut, and would essentially double the capacity of the secondary system without a comprehensive plant expansion.

There was initial doubt on the part of the regulatory agencies concerning the success of this project since a venture of this kind had not previously occurred in the State of Connecticut. After careful analysis of the project, all agencies and parties approved this plan and the facilities were constructed. This paper will discuss some of the unique features of this cooperative project. First, these facilities were generally designed to industrial standards. Second, the contractor was selected as a result of a public bid. Third, all the processing equipment was purchased by Frito-Lay. Fourth, the facilities were designed, constructed and placed into operation under the supervision of Frito-Lay. Fifth, no federal funds were involved. Sixth, the POTW expansion was financed be industrial revenue bonds.

Project Scoping & Negotiations

The town of Killingly, Connecticut is located in Windham County in the northeast corner of the state. In spring 1978, Frito-Lay selected Killingly as the site of one of its new plants. Killingly offered convenient access to the major Connecticut, Massachusetts, Rhode Island and New York markets, a solid work force, acceptable environmental costs, and strong interest by the town leadership in the project.

One of the more complex issues to resolve was treatment of Frito-Lay's 1 million gallon per day high strength wasteloads. A number of options were studied separately by Frito-Lay, and jointly with the Killingly Sewer Authority. Ultimately, a decision was made for Frito-Lay to construct a full primary treatment plant on-site, discharge the effluent to the sewer system, and build an expansion to the POTW. The back-up alternative was for Frito-Lay to build on-site secondary treatment and discharge to the sewer. This would be done if Frito-Lay and Killingly could not reach agreement, unexpected technical problems arose, or if costs became uncontrollable.

The Killingly Sewer Authority operates an 8 million gallon per day secondary treatment plant. The POTW was completed in 1976 with the intent of providing substantial reserve capacity in order to provide for future population growth and to attract industry. This plant has a raw sewage lift station, primary settling tanks, a conventional activated sludge system, secondary settling tanks, and chlorine contact tanks. Solids handling consists of thickening both primary and secondary solids prior to dewatering on rotary vacuum filters. Sludge is lime treated, then landfilled. An on-site sludge incinerator will be used as back-up, or when incineration of large sludge volumes becomes economical. Frito-Lay's discharge to the POTW would consist primarily of soluble BOD, and some solids. However, even with full primary treatment of Frito-Lay's wastewater, the POTW could not handle the wasteload. Secondary BOD was the problem. An expansion to the BOD capacity might provide adequate capacity. A study was initiated to investigate such an expansion.

If a full secondary treatment system were to be installed on-site by Frito-Lay, not only might it cost more than an expansion of the POTW but it would result in the duplication of equipment already installed at the POTW. In addition, Frito-Lay would have to cope with secondary sludge disposal. Costs were believed to be compatable or significantly lower, if the POTW was expanded. Therefore, Frito-Lay agreed to design and construct a secondary pump station, a biofilter tower, instruments and controls, sludge piping and related pumps, valves, necessary appurtenances, and modifications to appropriate existing components at the POTW.

The proposal to take an essentially new and under-utilized POTW, expand it, and change the process concepts was received with some skepticism by both the Killingly Sewer Authority and Frito-Lay's management. Major negotiating and paperwork nightmares were envisioned. Project delays would impact Frito-Lay's start-up.

Regulatory authorities were expected to question the project, even though no change in the NPDES permit limits or requests for federal funds were involved.

POTW Inprovements

Parallel to examining these procedural questions, the Killingly Sewer Authority, the town's consultants, and Frito-Lay's engineers began unit-by-unit analyses of the POTW. Numerous meetings and iterations were involved. The final result was a decision to expand the POTW by adding a pump station, a packed media trickling filter tower, and assorted controls. The POTW would be converted from activated sludge to a trickling filter/roughing filter process. The BOD capacity would be almost doubled. All other systems were analyzed to be adequate. Long term marginal capacity in clarifiers and flotation units might be a problem if the city and Frito-Lay both reached ultimate capacity. However, since this event is anticipated to be at least 10 years in the future, work on these systems was deferred.

The selected option called for converting the POTW from a conventional activated sludge system to a biofilter/activated sludge system. These improvements to the Killingly POTW almost doubled the BOD capacity of the secondary system (from 8,000 to 15,700 lb/day) and retained its current peak hydraulic capacity of 24 mgd. These benefits were achieved by constructing a minimum number of units. However, because the original plant was not designed with this type of expansion in mind, many of the connections and modifications were difficult and expensive.

The pump station utilizes five constant speed, submersible pumps to lift the primary effluent to the top of the biofilter tower. Capacity of the pump station varies 4,700 gpm with one pump running, to 17,000 gpm under peak flow conditions. This allows hydraulic wetting rate to be varied from 1.2 to 4.2 gpm/sf of surface area

of the biofilter. The pump station utilizes simple controls consisting of On/Off switches coupled to low-level and high-level sensors. One of the five pumps will be connected to the treatment plant's standby generator so that process integrity is maintained in the event of power outage. The number of pumps in operation will vary according to flow conditions, but under peak design flows only four pumps will operate with the remaining pump held in reserve to comply with State regulations.

The discharge manifold, gate valves and check valves are contained in a vault adjacent to the pump station. Access to this vault is via two large doors. The pump station was constructed of reinforced concrete. The wet well and effluent channel are open but are protected by handrails. An overhead electric hoist was provided to facilitate pump maintenance. All the pump starters and controls are located in a NEMA 14 enclosure installed between the pump station and the biotower.

The biofilter tower is approximately 70 feet in diameter and 28 feet tall. The biological growth media consists of horizontal redwood slats. The media is supported by pressure treated fir stringers which are supported by a concrete underdrain system. Media depth is 21.5 feet resulting in a net media volume of 80,000 cubic feet. The tower walls were constructed of precast concrete panels trimmed with face brick to match the existing plant structures. The design loading is 200 pounds BOD per 1000 cubic feet and 65% removal efficiency is projected. In sizing the tower, it was assumed that no additional BOD from Frito-Lay would be removed across the existing POTW primary treatment system.

Wastewater is uniformly distributed over the media by a four-arm rotating distributor. As the wastewater splashes through the biofilter, bio-solids (microbes and bacteria) form on the media and reduce the BOD of the primary effluent. Bio-tower effluent is returned via the

underdrain system to the pump station. A broad-crested weir divides this flow with a portion being recycled through the tower to "seed" the process at a typical internal recycle rate of 3:1 to 5:1. This serves to increase removal efficiency and dampen shock loads. The remainder of the bio-tower effluent flows by gravity to the aeration basin (existing) for further treatment.

In order to convert the Killingly POTW from an activated sludge to a biofilter/activated sludge system, substantial improvements, additions, and renovations were made to existing components. More than half the cost of the expansion project involved such "remodeling" items as:

- Removal and relocation of major slide gates, valves, and piping connections (30 and 36-inch).

- Removal of existing primary to aeration basin piping, and routing a 24" pipe to the new pump station. Several other pipes were rerouted to achieve this.

- Core bore gallery walls to allow for 36" piping extensions to the new pump station, and then reconstruct the walls.

- Route 36" return pipe from biofilter through existing pipe gallery to two aeration basin division boxes, including 36" valves and other miscellaneous connections. Stainless steel pipe was used to facilitate installation in very tight quarters where future maintenance would be difficult.

- Extend and reroute various sludge piping systems to the new pump station. Stainless steel pipe was also used here.

- Repair all damaged or modified basin and tunnel walls to be fully watertight.

- Upgrade electrical system to handle the increased motor load and install such conduit runs, starters, controls, etc. as were required for the new systems.

- Remove and extend existing flood plain retaining wall, drainage system, paving, fencing, and rip-rap to provide space to install biofilter and pump station.

- Place 5,000 cubic yards of select structural fill on which the biotower and pump station were constructed.

CONSTRUCTION OF THE EXPANSION

Since construction required difficult tie-ins and the POTW had to continue in operation, contractor selection was critical. During the design period, approximately 75 contractors were interviewed. From these, 25 contractors prequalified and were invited to bid. Many of the contractors were concerned about the risks of working for a private company (Frito-Lay) to construct facilities on public (Authority) property. Two actions were taken to address these risks. First, the specifications were written in the Construction Specification Institute format. They explicitly set-forth the General Conditions which related to this project, especially as to how the contractor was to interface with Authority personnel and Frito-Lay. In addition, the specifications contained the contract to be executed between the successful contractor and Frito-Lay. These details allowed many questions to be answered about the contract before bids were received.

Secondly, a comprehensive pre-bid meeting was conducted on-site. The plans, specifications, and site were thoroughly reviewed. All present reached agreement on the best way to handle specification addenda. Finally, it was announced that the contract would be awarded to the qualifying low bidder at the bid opening if all paperwork was in order and the bids contained no exceptions. There

would be no lengthy delays, "backroom negotiations" or pressure to lower bids. We hoped to eliminate contingencies, padding of the bids, and project delays.

These actions resulted in the receipt of 12 lump-sum bids. Four of them were under the engineer's estimate. One contractor was disqualified for failure to provide a bid bond. Within 30 minutes after the last bid was opened and qualified, the contract was awarded to the low bidder, R.H. White Construction Company of Auburn, Massachusetts.

Frito-Lay and Killingly were committed to making these facilities work as designed and on schedule. This required the pre-selection and pre-purchase, and expedited delivery of all the process equipment. Frito-Lay directly purchased the biotower media, rotary distributor, submersible pumps and controls, all valves over 12 inches, and the flow meter and instrumentation. The detailed design and specifications were completed after these purchases were made so that more exact details were worked into the plans. There would be no "or equals" or substitutions that might cost more to install, lead to contractor bid contingencies, or not meet the process requirements. As a result the contractor knew before submitting a bid exactly what equipment had to be installed and when it would arrive on-site. Finally, all construction was supervised under the direction of Frito-Lay, utilizing the field engineer on-site at the production plant (4 miles away) and a representative of the consulting engineer.

These actions resulted in a high quality project that was completed on time and within budget. Killingly Sewer Authority personnel participated in every phase of the process design, equipment specification and selection, and in construction coordination. Their involvement from an operating standpoint was vital. Although designed to industrial standards, the facilities augment the architecture of the POTW and they are completely acceptable to the Authority.

PROJECT FINANCING

The final unusual feature of this project is the method of financing the improvements. Due to funding priorities and the industrial nature of the POTW capacity expansion, federal funds were not available. A decision was made to obtain pollution control industrial revenue bonds (IRB's) through the Connecticut Development Authority. In total, $6 million in bonds were obtained to pay for various pollution expenditures associated with the new plant project. The $2 million allocated to the POTW expansion could only be approved if Frito-Lay, Inc. continued to own the improvements for the twenty-five year life of the bonds. After considerable negotiations and legal opinions, a contract was executed that allowed Frito-Lay to own the facilities and depreciate them. The Killingly Sewer Authority is totally responsible for oeration maintenance and repair. By utilizing IRB's, Frito-Lay reduced the annual interest rate to 6.3% from the 10-11% rate prevailing at the time of the closing on the bonds. Abailability of IRB financing encouraged Frito-Lay to proceed forward with the POTW expansion program in Killingly.

SYSTEM PERFORMANCE

The trickling filter was placed in operation during April, 1981. The biotower acclimated quickly, even though it was started in the roughing filter mode with no seed addition. In mid-August, the decision to convert to the "ABF" mode was made in order to obtain warm weather operating data before the onset of winter. A license to operate the Activated Bio-Filter (ABF) process was issued by Neptune - Microfloc, Inc. when the redwood media was puchased. Table 1 summarizes system performance to date.

The roughing filter performed as expected. However, the biotower was designed to operate in the ABF mode. Starting in August, the removal efficiencies increased dramatically from 65 percent to a high of 75 percent in January, 1982. Also in January, the removal rate reached 52.8 pounds BOD per 1000 cubic feet. Average wastewater

temperature declined from 21.0°C in August to 11.5°C in January.

Table 1. Summary of System Performance

Month	Loading Rate (lb BOD/1000 CF)	Removal Rate (lb BOD/1000 CF)	Removal Efficiency (Percent)
June, 1981	47.6	24.9	52.4
July	48.7	22.8	46.8
August	46.8	30.9	65.9
September	50.4	33.7	66.9
October	72.1	47.1	65.4
November	42.5	27.1	63.8
December	59.1	41.3	69.8
January, 1982	70.3	52.8	75.2

During the period August through November, biotower internal recycle was kept to a minimum and the return activated sludge rate applied to the tower was 100% of plant influent flow.

Shortly after start-up, there was a marked improvement in the effluent suspended solids discharged by the POTW. In addition, solids thickening operations improved. Most importantly, sludge dewatering costs decreased from a high of $57.33 per ton to a low of $19.30 per ton. Table 2 summarizes vacuum filtration chemical conditioning costs during the study period.

Table 2. Vacuum Filtration Chemical Costs

PERIOD	DRY TONS DEWATERED	AVERAGE COST/TON
January through April, 1981	329	$ 43.15
May through August, 1981	385	27.06
September through January, 1982	471	28.97

During this time, there was a 15 percent increase in the quantity of dry solids dewatered but a 33 percent decrease in the chemical conditioning costs per ton of dry solids. These cost savings amount to $15,500 annually. Less time is required to process these solids. Labor and power savings are also substantial.

Conclusion

Based on operation to date, the biotower is achieving the design removal efficiency. Overall, the POTW operation has improved. Average effluent TSS and BOD values are lower now than before Frito-Lay started production. There have been few permit violations, these being minor suspended solids excursions above the 30 mg/l level. Settling characteristics of the secondary solids have improved. In fact, it costs substantially less now to dewater sludge than two years ago in spite of inflation and even though sludge quantities have increased substantially.

The Authority is sufficiently pleased with the biotower facility that Frito-Lay, Inc. was able to negotiate an expansion of production. In fact, this expansion is currently underway and will make the Killingly plant Frito-Lay's largest.

THE EVALUATION OF A BIOLOGICAL TOWER FOR TREATING AQUACULTURE WASTEWATER FOR REUSE

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INTRODUCTION

During the 1950's, Dr. Shao-wen Ling was sent to Southeast Asia by the Food and Agriculture Organization (FAO) as a fisheries specialist. He became interested in a large Malaysian prawn which he first observed in the marketplace. Since that time there have been many species of freshwater prawns identified that are distributed over the tropical and semi-tropical waters of the world. Most are limited to waters that maintain annual temperatures of 22 to 30 C. The most popular species for culture is Macrobrachium rosenbergii (1).

Adult prawns are capable of living in freshwater and may be either intensively cultured in tanks or raceways at high stocking density or extensively cultured in earthen ponds at a lower density. Marketable size prawns may be harvested in five to six months under optimal conditions. Glude (2) presented an overview of freshwater prawn culture that describes rearing requirements and technques. Figure 1 illustrates the adult prawn showing characteristic parts.

The development of prawn larvae to metamorphasis requires temperatures of between 24 and 32 C with preferred temperatures around 28 C. These temperatures also apply for commercial growth of adults in ponds.

Adults and broodstock may be maintained easily in either freshwater or brackish water. The larvae require brackish water of between 8 and 17 ppt salinity. Juveniles require salinities of about 5 to 8 ppt decreasing to freshwater as they grow to adults.

The toxicity of ammonia to aquatic life has been fairly well documented. The EPA criterion of unionized ammonia for fish and aquatic life is less than 0.02 mg/l (3). The 144 hour LC 50 for Macrobrachium rosenbergii was determined by Armstrong (4), to be 0.80 mg/l ammonia at a pH of 7.6. Additional work is necessary to verify and establish these toxic limits of unionized ammonia for the Malaysian prawn.

Nitrite has also been shown to be toxic to freshwater prawns (5). LC 50 values ranged from 500 mg/l for the first 12-hour exposure to 5 mg/l during a 168-hour exposure. The maximum level of nitrite tested with no deaths, ranged from 9.7 mg/l for 24 hours to 1.8 mg/l at 168 hours.

There is a paucity of data concerning the concentration and nature of wastes produced in intensively cultured systems. Little has been presented about water quality of effluents from culture systems. The problem is complexed by the fact that organics and nitrogen in the water may be due to metabolic products of prawns and other secondary feeders in the culture ponds as well as unused feed added daily to the water. A summary of the reported water quality values and the proposed levels for design of new production facilities is shown in Table 1.

Several investigators have described water recirculating systems using biofilters for intensive cultivation of salmon smolts (6,7,8), catfish (9), trout (10,11,12,13), shrimp (14,15,16), tilapia (17,18), polyculture (19), carp (20), and the combination of fish culture with hydroponics in a single recirculation system (21). Few of the authors, however, discuss the basis for their choice of biofilter design parameters. There is little indication that the designs conform either to economic or to resource minima for their declared purposes. Spotte recommends that surface area of the filter be made at least equal to the surface area of the culture. This rule of thumb may favor successful recirculation water quality but says little about keeping the investment minimal (22).

Removal of the toxic hazard from nitrogenous waste is the most important task of the biofilter. The predominant form of waste nitrogen from aquatic animals is ammonia, excreted mostly from the gills or in the urine, or produced by mineralization of organic nitrogenous substances by bacteria. Invertebrates also excrete nitrogen in the form of ammonia. Bacteria of the Nitrosomonas group oxidize ammonia to nitrate;

$$NH_4^+ + OH^- + 1.5\ O_2 = 2\ H^+ + NO_2^- + H_2O$$

the reaction is rate limiting in the overall process of nitrogen control. For this reason, nitrites seldom accumulate in biofiltered recirculation water. Nitrite is readily oxidized to nitrate by Nitrobacter ssp.;

$$NO_2^- + 1/2\ O_2 = NO_3^-$$

and the nitrate is then available either for assimilation by green plants or bacteria, or for reduction to nitrous oxide and free nitrogen, usually by anaerobic denitrifying bacteria.

The microbial population activity in the filter is essentially ubiquitous in distribution, spontaneously coloniing new filters, although inoculation can be expedited by addition of gravel from preexisting filters (8). The bacterial species which predominate at various depths of the filter are presumably self-selecting, in response to environmental factors, such as the nutrient content, hydraulic flow, and aeration of the filter bed, although their respective numbers are known to fluctuate until an equilibrium is established (Kawai, et al., 1964).

Few approaches to analysis of design standards for biofilters in aquacultural applications are given in the literature. Hirayama reported on filter carrying capacity for a 300 liter marine aquarium containing one or two sea breams, Chrysophrys major (23). Spotte cautions that Himayama's results need verification before accepting their validity in freshwater, but there is reason to exercise a certain amount of skepticism in generalizing them even for marine aquaria (22). One should keep in mind the small numbers of fish on which they are based.

Harris considered use of biological filters in hatchery water reuse systems. Submerged filters (Flexring and Flocor) were tested and found satisfactory in control of fish wastes. The basis for design of the biofilters used was Speece's method for finding substrate volume based on ammonia production rate divided by the nitrification rate (10).

A major limiting factor in intensification of aquaculture is water quality. Biofilters need to be included in aquaculture systems to insure pollution removal and good quality water for use and reuse. Work is being done on optimizing treatment systems for aquatic animals and developing accurate design specifications for treatment. More research is needed in this area.

Management of the filter is an important consideration in obtaining nitrification and BOD removal from either fresh or saline aquaculture wastewater. Removal rates may be enhanced by increased temperature, recirculating effluent back over the filter media, dosing the filter, ventilating the filter, adding additional media and surface area, and maintaining environmental requirements by careful monitoring.

Wheaton (24) describes in depth design criteria for submerged filters with varied filter media. The equations presented for design are restricted to cold-water applications and care is advised in extending to treatment systems for treating wastewater produced in warmwater fish culture facilities.

The types of media that have been applied to aquaculture include oyster shell, rock and gravel, plastic rings, poly beads, sand, and styrofoam. A new possibility is the Tri-Pack spherical media available from Jaeger Tri-Packs in Costa Mesa, Calif (technical bulletin, 1981).

Trickling filters and upflow submerged filters may be designed according to a simplified procedure presented by Soderberg and Quigley (25). Their data is for perch culture and probably should not be generalized for culture of all warmwater species of fish and invertebrates.

This study is a joint effort between Dr. Stanley L. Klemetson currently at Brigham Young University and Dr. Dan Cohen of the Hebrew University. The U.S. portion of the study deals mostly with engineering problems while the Israel portion deals with production of the animals. It has been important to coordinate efforts and share information to improve each study.

METHODOLOGY

The filters used in this study were laboratory-scale units having a capacity of 1.4 cu. ft. each. Each filters was filled with 1 to 1.5 inch slag with a total surface area of 82.0 sq.ft., and a specific surface area of 60.0 sq.ft./cu.ft.

Figure 2 presents a schematic of the two filters used in this study. The facilities at Brigham Young University were used to construct the columns. Both filters were fabricated from a section of 10 inch PVC pipe. The end discs were cut from plexiglass.

The two filters were fed a uniform flow from a 4.5 cu. ft. plexiglass constant head tank with overflow weir (Figure 3). Two 4.0 ft (0.11 m) storage tanks filled with synthetic wastewater was used to feed the constant head tank.

The two storage tanks were used to hold the synthetic wastewater to be treated by the system. The synthetic wastewater was compounded as needed from a balanced minimal media which approximates actual wastewater produced by prawns. Table 2 presents the composition of the synthetic feedstock solution. Diffused air was introduced to insure adequate mixing within the storage tanks. Tap water was used to fill the tank to its desired level.

The studies done at the Hebrew University in Israel utilized 900-liter aquaria and submerged filters. The submerged filters were filled with 20-30 cm gravel in a 700 l tank. The flowrate was controlled at 0.5 cu. m./ hr. Salinity of the rearing tank was maintained at 1.2% sea water for growth of freshwater prawn larvae.

Water quality sampling was done prior to feeding and drainage of sediments for the Israel study. Standard Methods (26) were used to determine dissolved oxygen, biochemical oxygen demand, pH, alkalinity, salinity, ammonia, nitrate,and nitrite both at the Hebrew University and at Brigham Young University. Water samples were taken at both the influent and effluent of the biofilters.

RESULTS

The results of work done at Brigham Young University are presented in Table 3. After acclimation of the biofilters, data was collected to evaluate their performance. Profiles of the parameters monitored are presented in Figures 4

through 7 for the Submerged and Trickling filters. Figure 4 presents a profile of ammonia levels. The average value of ammonia removal for the trickling filter was 78%. In the case of the submerged anaerobic filter, there was an average increase of ammonia by 8% through the filter.

Figure 5 presents the nitrate profile for the two filters. The nitrate concentrations decreased by 97% in the submerged filter while there was an increase observed in the trickling filter. The length of time required for start-up of the filters was 3-4 weeks. At three weeks, the trickling filter was effectively nitrifying the synthetic wastewater. The submerged filter was anaerobic and functioning as a denitrifying filter in about the same length of time. Other parameter profiles are presented in Figures 6 and 7.

Data describing the results of the Israel study are presented in Table 4 and 5. The data indicates that the nitrification efficiency of freshwater prawn wastewater was 59% ammonia removal and 55% nitrite removal. Figure 8 presents the nitrogen profiles for a tank and biofilter system in Israel. The filters were capable of maintaining a stable environment, though further study is necessary to determine sizing criteria for filter design.

The results of these studies are being reviewed as part of an ongoing study to develop design criteria for application to treatment of warmwater aquaculture wastewater. The data obtained in this portion of the study suggests that nitrifiction efficiency can approach 80% when the synthetic wastewater is lightly loaded. The optimum hydraulic loading has not been determined. The efficiency of ammonia removal in brackish water is about 60%.

An important consideration in maintenance of the biofilter is aeration. If aquaculturists are interested only in nitrification, careful monitoring and addition of dissolved oxygen is important. Denitrification is an anaerobic process that requires little maintenance. Aquaculturists should consider the coupling of the two processes, especially where the recontioned water is returned to the culture vessel.

DISCUSSION

The design and sizing of biofilters for application to aquaculture is more of an art than a science. If facilities are to be developed for intensification of fish culture, a great deal of research needs to be done. Design criteria are

available for facilities that rear cold-water fish and invertebrates but is limited to water temperatures up to 15 C. Since operating temperatures are high (27 to 30 C), organic loadings are small, and flowrates are high, standard design tables are not adequate. To complicate matters more, the actual waste production rates for the live aquaculture systems are variable and ill-defined. Optimization of design will require the continued evaluation of equipment and processes under a variety of conditions.

ACKNOWLEDGEMENT

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Figure 1. Freshwater Prawns (Macrobrachium rosenbergii) and Artificial Habitats in Fish Culture Tank.

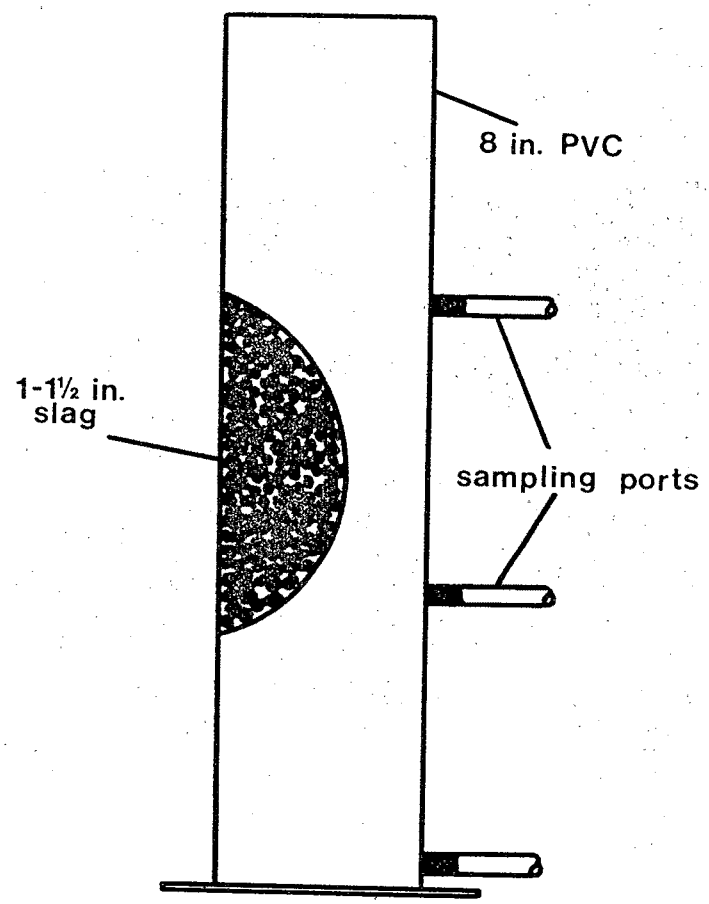

Figure 2. Schematic Diagram of Biological Tower and Submerged Filter (BYU Study).

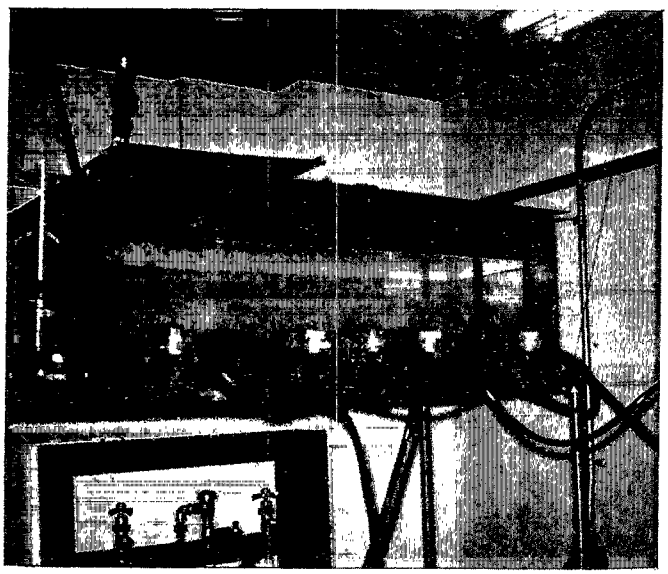

Figure 3. Constant Head Tank Used to Dose Biological Tower and Submerged Filter (BYU Study).

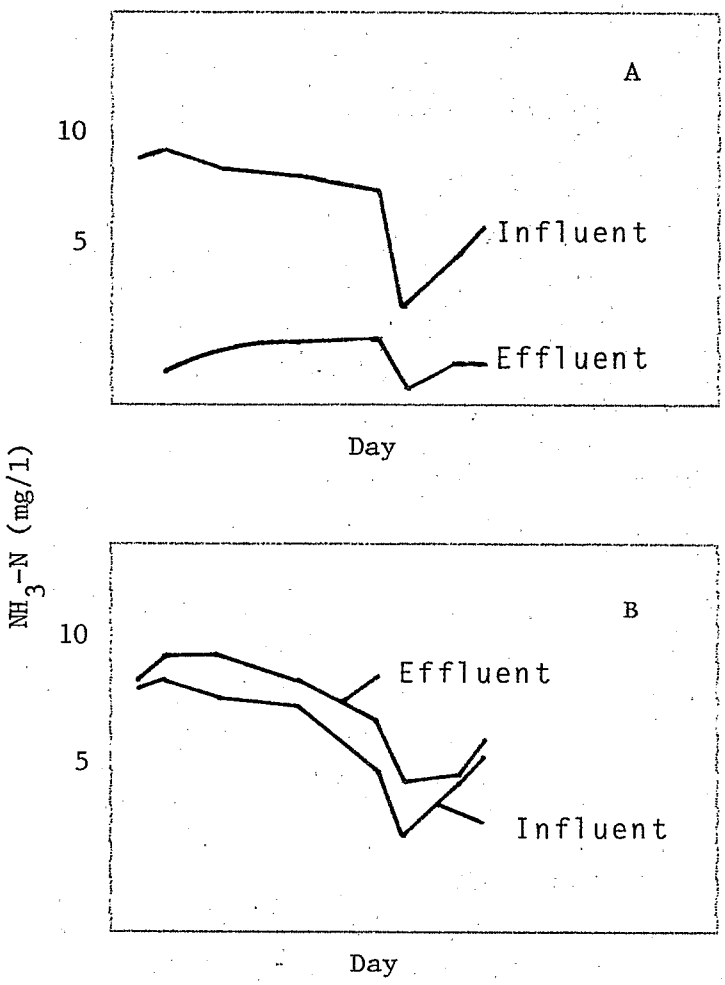

Figure 4. Ammonia Removals in Biological Tower (A) and Submerged Biofilter (B).

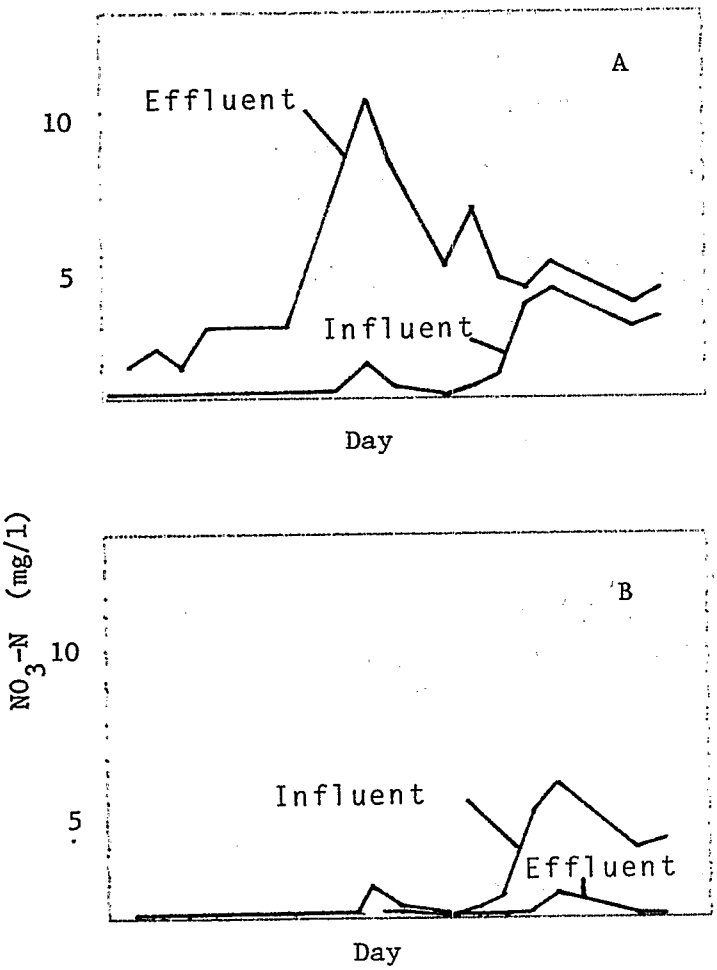

Figure 5. Nitrate Removals in Biological Tower (A) and Submerged Biofilter (B).

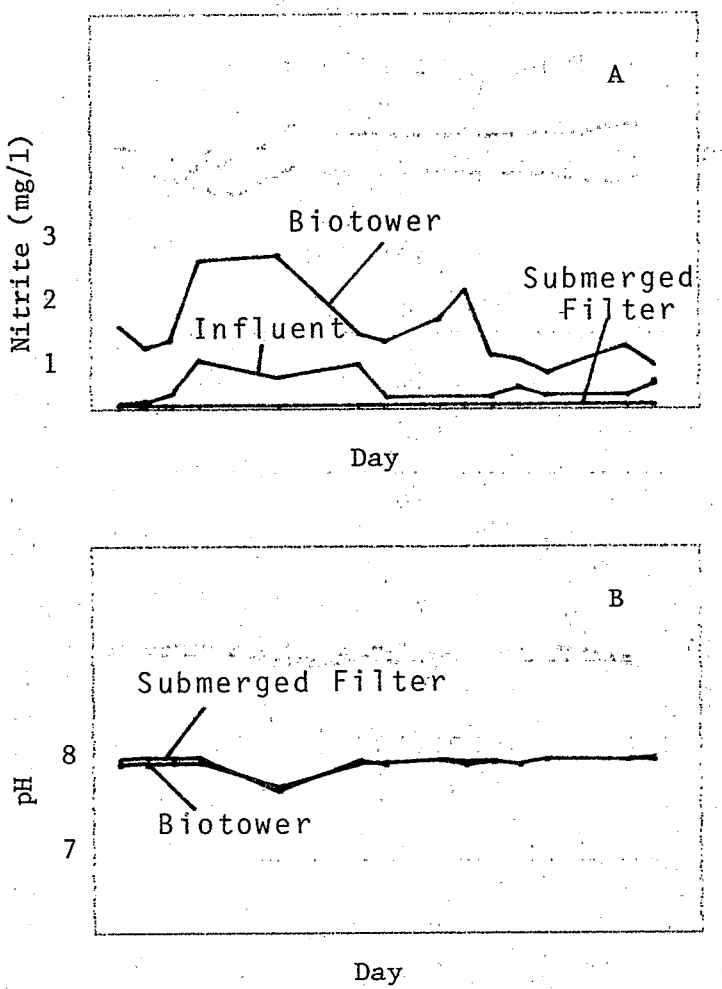

Figure 6. Nitrite (A) and pH (B) profiles for Biological Tower and Submerged Filter.

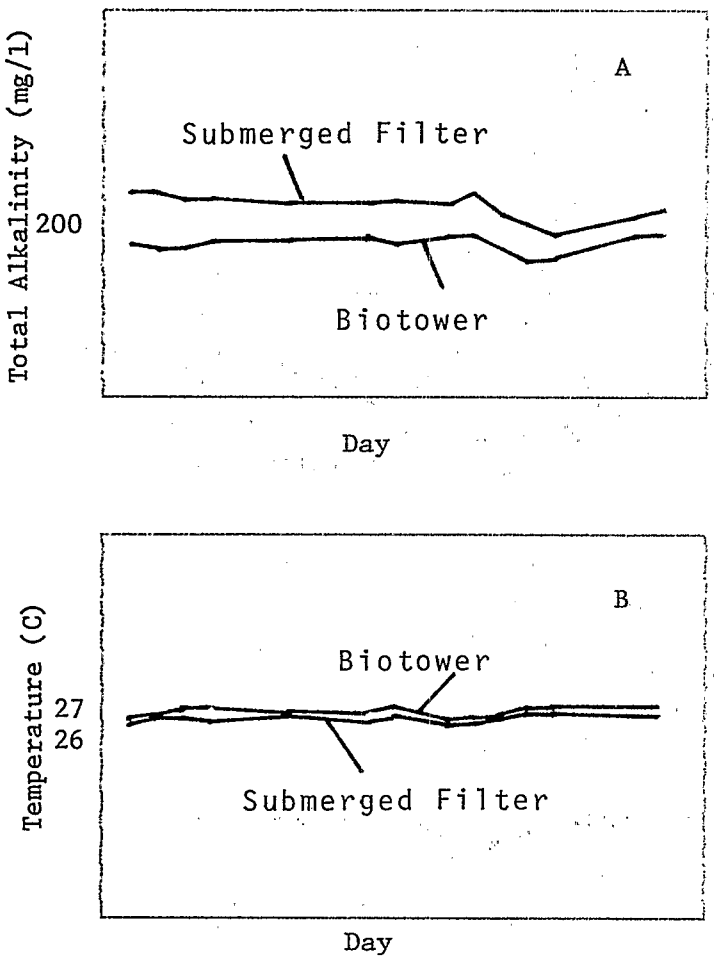

Figure 7. Alkalinity (A) and Temperature (B) Profiles For Biological Tower and Submerged Filter.

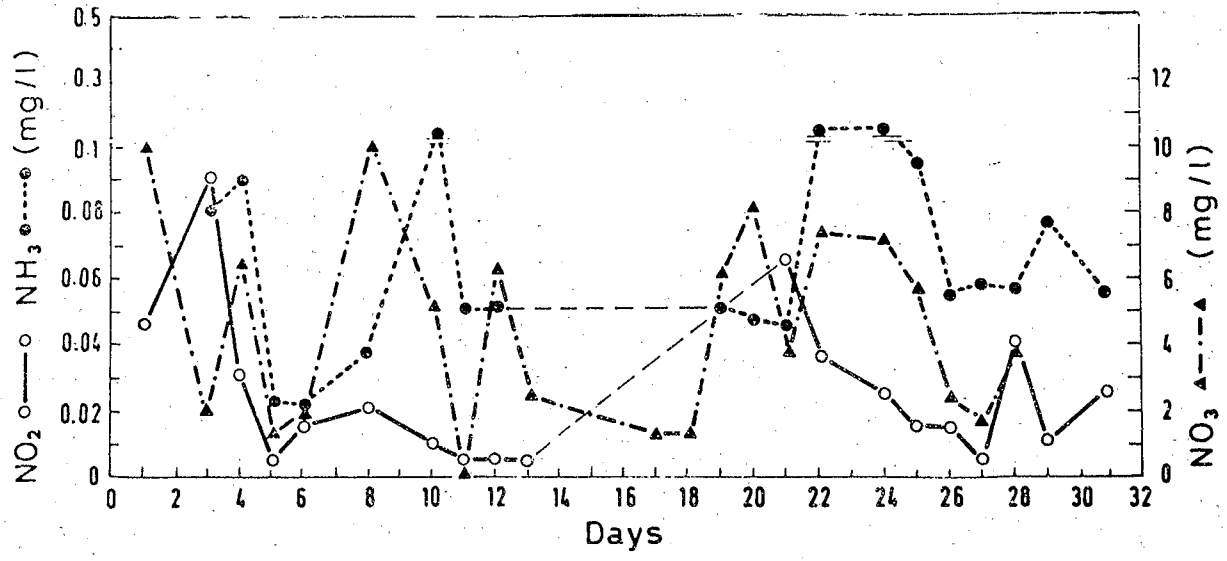

Figure 8. Nitrogen Profile for Recirculating Prawn Hatchery (Israel).

Table 1. Reported Values of Water Quality for Aquaculture Ponds.[a]

parameter	US EPA[b]	Roberts[c]	Klontz[d]	Nightingale[e]	Canada[f]	Suggested Level
Aluminum				0.2	0.1	<0.1
Ammonia	0.02	0.02	0.012	0.1	0.02	<0.02
Cadmium	1.2-12	0.4-3.0		0.05	0.2	0.2-12
Calcium			52			>52
Carbon Dioxide			2.0			<2.0
Chromium	0.1	0.05		0.5	0.04	<0.1
Coliform	14					<14
Color Units	75					<75
Copper	1.0	0.01	0.006	0.02	0.005	<0.01
Dissolved oxygen			5.0		4.0	>4.0
Iron	1.0		1.0	0.5	0.3	<0.5
Hardness				300		<300
Lead		0.03		0.1	0.03	<0.03
Magnesium						
Mercury		0.05			0.1	<0.1
Manganese	100					<100
Nitrate						
Nitrite		0.1	0.55			<0.1
pH (pH units)	6.5-9.0	6.5-8.5	6.7-9.0	6.5-8.0	6.5-9.0	6.5-9.0
Sulfate						
Sulfide	0.002	0.002	0.002		0.002	<0.002
Specific Conductance						
Temperature				50-90°F		50-90°F
Total Alkalinity	20		20-200		20	20-200
Total Dissolved Gas		110%	110%	105%		<105%
Total Dissolved Solids	250		400			<400
Total Suspended Solids		8	80		25	<80
Turbidity (JTU)				60		<60
Zinc			0.04	0.1	0.03	<0.03

[a]All concentrations in mg/l except as noted
[b]US EPA Redbook, 1976, Water Quality Criteria
[c]Roberts, Fish Pathology
[d]Klontz, et al., A Good Deal From Egg Sac to Creel,1979
[e]Nightingale, Development of Biological Design Criteria For Intensive Culture, 1976
[f]Environment Canada, Water Quality Sourcebook, 1979

Table 2. Synthetic Wastewater Solution (BYU study).

Dextrose	100 mg/L
Yeast Extract	10 mg/L
Urea	5 mg/L
Na_2HPO_4	40 mg/L
$NaHCO_3$	125 mg/L
$MnSO_4$	2 mg/L
NH_4Cl	as required

Table 3. Ranges and means of Water Quality Data (BYU Study).

	NH_3-N [a]	NO_2-N [a]	NO_3-N [a]	pH [b]	DO [a]	Alk [a]
Influent	3.7-9.6 (7.61)	0-3.57 (0.89)	0-5.19 (1.28)	7.5-8.2 (7.88)	1.0-6.2 (3.88)	166.9-233.9 (207.46)
Trickling Filter Effluent	0.52-2.5 (1.64)	0.5-2.68 (1.30)	1.1-11.53 (4.47)	7.4-8.1 (7.81)	3.7-6.3 (5.30)	156.2-210.0 (178.75)
Submerged Filter Effluent	5.8-10.7 (8.25)	.001-0.28 (0.03)	0.01-0.96 (0.12)	7.8-8.1 (7.89)	0.2-1.4 (0.97)	187.6-236.8 (219.94)

[a] data reported in mg/1

[b] data reported in pH units

Table 4. Ranges and Means of Prawn Water Quality Data (Israel Study).

TEMP. MAX.	MIN.	D.O.	pH	NH_3-N	NO_2-N	NO_3-N
30 (28)	26	6 - 9.25 (7.445)	7.7-8.2 (8.082)	0.7-0.038 (0.219)	0.570-0.005 (0.0715)	10.0-0.05 (4.58)
32 (26)	15	6.5-8.25 (7.158)	7.85-8.50 (8.276)	0.28-0.005 (0.093)	0.03-0.002 (0.009)	9.8-0.05 (4.25)
35 (30)	29	5.8-9.0 (7.136)	7.98-8.30 (8.153)	0.235-0.035 (0.099)	0.075-0.005 (0.02)	10-0.05 (4.5)
33 (29)	28	5.0-8.5 (6.9)	8.075-8.330 (8.168)	0.190-0.025 (0.087)	0.085-0.005 (0.027)	9.8-0.05 (4.836)
31 (29)	27	5.52-8.25 (7.160)	7.975-8.370 (8.130)	0.230-0.035 (0.094)	0.150-0.005 (0.032)	10.5-0.05 (4.186)
32 (28)	20	5.0-8.10 (6.68)	8.45-8.325 (8.17)	0.150-0.020 (0.069)	0.090-0.005 (0.025)	10-0.05 (4.47)

Table 5. Removal Efficiencies of Biofilters (Israel Study).

TANK NO.	NH_3-N	NO_2-N	NO_3-N
3	36	54	33.
5	52	42	31.
9	58	50	27
10	70	53.	35.
11	73	69	12.
12	64	62	29
$\bar{x}$	58.8 ± 13.6	55 ± 9.3	28 ± 8.5

BIOFILTRATION OF TANNERY WASTEWATER

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INTRODUCTION

Industrial pollution on an unprecedented scale has emerged as one of the most pressing problems throughout the world. While governments have been grappling with a pernicious combination of economic, social and political problems, they have not paid equal attention to issues related to environmental protection. Improving the quality of the environment is no longer a luxury measure; it is one which, in the long run, will generate immeasurable benefits in terms of protecting public health and natural resources, and indirectly contributing to economic growth.

Alexandria is the principal port of Egypt, the country's largest industrial center and its prime resort. The ever-increasing discharge of heavily polluted industrial effluents from tanneries of the Mex Industrial Complex (MIC) into the Mediterranean has had an adverse effect on public health, fish production, navigation and the environmental quality of the area. The combined effluent of MIC averages 2 million cubic meters annually, with an estimated population equivalent of 400,000.

Biofiltration has been widely recognized as a reliable treatment process, which suits the needs of small to medium-size industries, due to its versatility, ability to take shock loads and relative ease of operation. Decreased popularity of biofiltration in comparison to other treatment processes has been attributed to inability of existing installations to meet emission limitations and the ineffectiveness of biofilters for treatment of concentrated industrial effluents.

Renewed interest in biofiltration in Egypt is due to relatively low power requirements, system flexibility, and the need for considerably less technical know-how for effective operation compared to other treatment technologies.

High-rate biofiltration was shown by Smith and Kates[1] to be cost-effective and capable of reducing organic loading. Several plastic media were tested in the high-rate biofilter units and each medium supported an adequate level of microbial organisms.

Problems with odor were encountered because of the sludge production of the high organic strength wastewaters.

Hosono and Kubota[2] reported that the BOD removal rate per unit of power consumption was shown to decrease with increasing BOD loadings. High-rate filters gave higher power economy values at higher BOD loadings, whereas standard-rate trickling filters were limited to low BOD loadings.

Bailey et al[3] have shown that high rate biofiltration can relieve overloaded conventional filters by removing about 40% of BOD from leather processing wastewater. Plastic media were used in the roughing filters which were dosed with tannery effluents containing vegetable tanneries at rates of 3.3 to 6.9 lb BOD/yd^3/d.

Pierce[4] reported that removal of BOD at two-stage filter plants is significantly higher than at single-stage plants. Chemical treatment with metal salts and polymers upgraded single-stage filter effluents from an average of 36 mg/l BOD to 21 mg/l. Similarly, effluent suspended solids were reduced from 32 mg/l to 19 mg/l. The cost of chemicals is not prohibitively expensive.

Previous research has indicated the need to assess high-rate biofiltration at various loadings and flow patterns for treatment of specific industrial wastes, in order to accurately evaluate system performance. It is particularly important to compare the performance of biofilters with other competing biological processes for treatment of tannery wastes and evaluation of cost-effective modifications to improve effluent quality. Recognition of these needs prompted the undertaking of this study.

BACKGROUND

The MIC tanneries are located east and west of the municipal slaughterhouse, as shown in Figure 1. At present, the slaughterhouse and tannery wastes (including organic particles and toxic chemicals) are collected in public sewers and discharged directly into the western harbour through three separate outfalls, without pre-treatment. Sewer clogging is frequently experienced due to large residues discharged with tannery effluents.

The General Organization for Industrialization (GOFI) is placing MIC tanneries on the top priority list of the most polluting industries which require Government technical and financial support for installation of waste treatment facilities. The available options being studied are: (a) primary treatment of combined effluent before discharge into public sewers for further treatment with domestic wastes or (b) biological treatment to meet Egyptian effluent limitations for direct disposal into water bodies.

According to prevalent practices in MIC tanneries, about 28-36 cubic meters of wastewater are generated per ton of hide processed[5].

Studies performed on the six major tanneries during 1980-1981 indicated that clean water pools of the beam house contribute 24.8% of the total effluent and only 0.28% of the BOD load, while vegetable tannery generates 1.4% of the liquid wastes and 43.3% of the BOD load (Table I). This suggests that judicious segregation of relatively clean process waters may appreciably reduce the size and costs of treatment facilities. The pollutional loads of tannery processes shown in Table II indicate that both beam house and tan-yard generate higher loadings than those originated from retin, color and fat liquor processes.

Chrometan mixed wastes comply with EPA guidelines[6] while BOD, COD and Oil and Grease (O & G) loadings of the vegetable tan mixed wastes were higher than those suggested in the guidelines. A summary of the physico-chemical characteristics and trace metal constituents of various process effluents are shown in Tables III and IV respectively.

MATERIALS AND METHODS

The experimental system consists of two biofilters, recycling pumps and clarifiers, as shown in Figure 2. Each filter

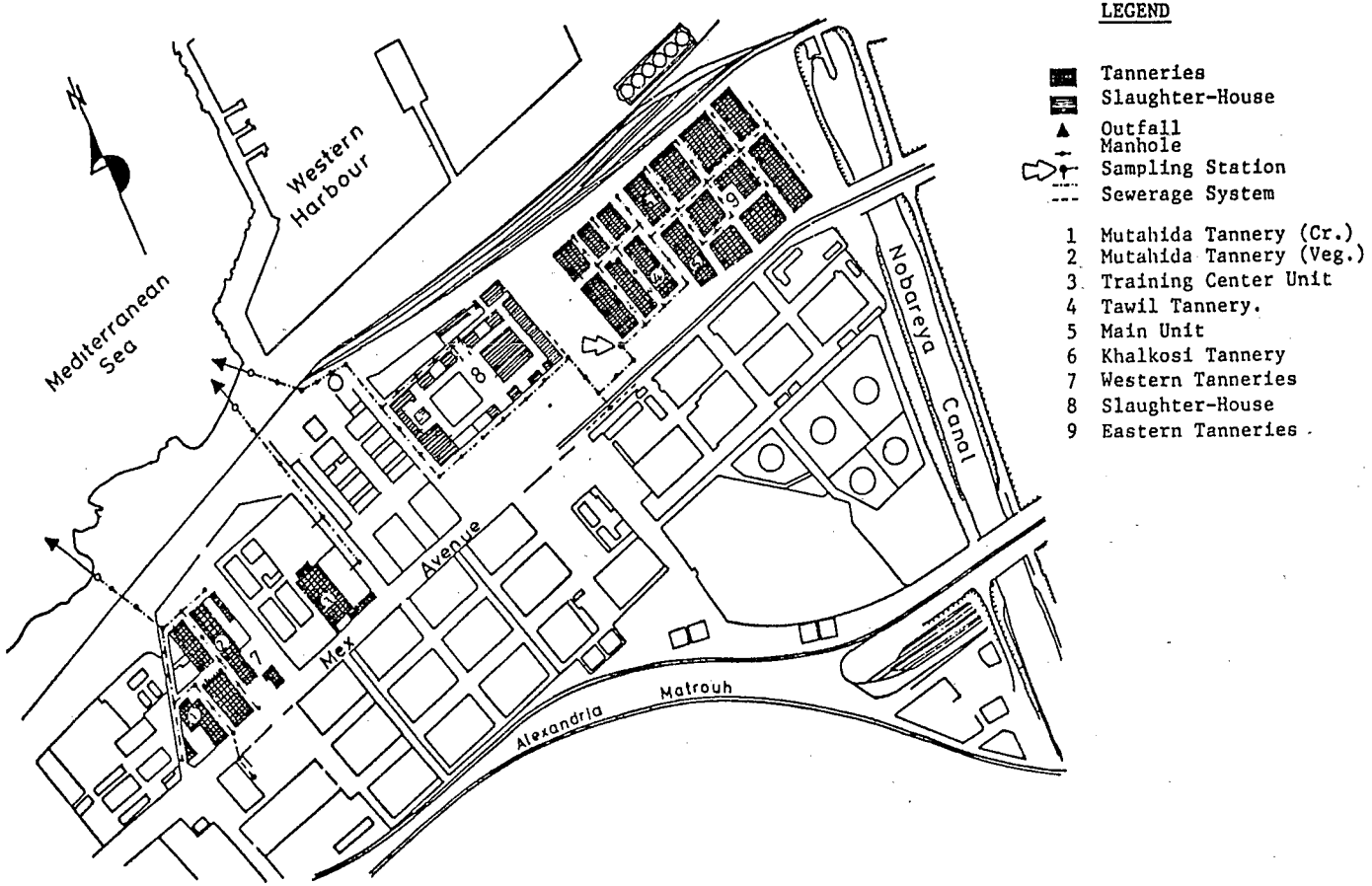


Figure 1. Site Location of MIC Tanneries.

Table I. Distribution of Volume and BOD load for the Processing Operations of The MIC Tanneries.

Process Waste	Volume % of Total Volume	BOD % of Total BOD Load
A. Beam-House Wastes		
Soaking & Washing	8.3	5.4
Lime & Unhairing	24.1	12.4
Delime & Baiting	20.0	7.7
Clean Wash Water Pools	24.8	0.2
	77.2	25.7
B. Tan - Yard Wastes		
Pickling	5.5	2.1
Chrome Tanning	5.5	3.4
Pre - Tanning	2.0	9.9
Vegetable Tanning	1.4	43.3
	14.4	58.7
C. Retan - Color & Fatliquor		
Neutralization	2.8	1.4
Bleaching	2.8	8.7
Color & Fatliquor	2.8	5.5
	8.4	15.6
Total	100	100

Table II. Pollutional Loads of the MIC Tanning Processes.

Effluent Stream	Pollutional Load Kg/ K Kg * of hides					
	BOD	TOC	TKN	SR	O & G	T. Cr
A. Beam-House Wastes						
Soaking & Washing	11.8	5.4	0.74	5.9	2.5	
Lime & Unhairing	26.6	15.5	1.64	51.8	2.2	
Delime & Baiting	16.5	8.2	5.3	27.7	0.4	
Clean Wash Water Pools	0.4					
B. Tan - Yard Wastes						
Pickling	4.5	2.3	1.6			
Chrome Tanning	7.3	3.7	0.7	9.1	0.1	
Pre - Tanning	21.2	10.6	0.75	9.5	0.6	
Vegetable Tanning	92.8		0.6	116	0.4	
C. Retan - Color & Fatliquor						
Neutralization	3.0	2.0	0.9	8.5	0.05	
Bleaching	18.7	15.1	0.5	51.2	0.3	
Color & Fatliquor	11.8	7.3	1.2	6.8	60.5	
Chrome-Tan Mixed Waste	92.6			109.2	8.6	3.5
Vegetable-Tan Mixed Waste	150.8			72.1	9.4	0.0
EPA (Ref.6)	95			140	19	4.3

Mean of 12 observations.

* K/KKg = Kg/1000 Kg.

Table III :Physico-Chemical Characteristics of the Wastewater Originating from MIC Tanneries

Process		pH (R)	Turb NTU	SR mg/l	Set.R ml/l/h	BOD mg/l	COD mg/l	TOC mg/l	NH_3-N mg/l	TKN mg/l	Cl^- mg/l	SO_4^{2-} mg/l	H_2S mg/l	Phenol mg/l	O&G mg/l
SOAKING	$\bar{X}$	7.4-	293	1669	107	3257	6370	1494	153	205	18460	1908	7.3	1.1	692
	SD	8.2	288	1169	14	2253	4420	1166	223	303	7481	1925	12.7	0.6	375
LIMING AND	$\bar{X}$	11.7	632	14359	506	7370	9892	4280	351	455	9060	3480	806	2.5	609
UNHAIRING	SD	12.5	114	4949	122	2902	4889	1842	376	350	2503	1256	683	1.9	185
DELIMIIG&	$\bar{X}$	8.5-	286	7703	233	4590	8141	2266	998	1466	10800	2659	39	14	108
BATING	SD	10.6	156	4947	91	1528	2933	1415	431	767	1673	2545	35	7.3	68
PICKLING	$\bar{X}$	2.8-	113	11082	139	1239	3391	637	295	441	17380	6460	1	0.2	75
	SD	5.1	129	22732	93	266	2171	106	211	335	7491	2497	1.1	0.2	54
CHROME	$\bar{X}$	3.1-	300	2530	179	2039	4460	1003	96	191	8850	4920	0.2	0.3	32
TANNING	SD	4.3	216	2616	83	938	1616	685	107	133	947	2667	0.3		28.4
PRETANNING	$\bar{X}$	4.1-	225	2590	257	5867	22747	2950	190	207	4633	3267	0.03	0.3	153
	SD	5.9	49	1566	51	4492	24864	636	93	99	785	1137	0.1	0.2	54
VEGETABLE	$\bar{X}$	5.2-		32272	227	35767	122267	28900	79	165	13233			3.3	101
TANNING	SD	6		21413	168	21435	30715	33102	20	58	8348			1.3	21
NEUTRALIZAT	$\bar{X}$	6.3-	100	2365	113	827	1679	567	177	248	3567	2107			14
- ION	SD	7.6	18	585	21	257	645	235	35	15	208	2255			5.4
COLOR &	$\bar{X}$	3.4-	37	1924	54	3390	5413	2011	132	336	3675	2933		0.1	1685
FATLIQUOR	SD	5.9	26	2950	19	2940	4942	1693	96	247	1742	686		8	1586

(R) = Range. $\bar{X}$ = Mean of four observations. SD = Standard deviation .

Table IV :Trace Metal Analyses of Wastewater from Processing Operations at MIC Tanneries.

Process		Total Chromium mg/l	TRACE METALS (ug/l)					
			Pb	Cu	Fe	Ni	Cd	Zn
Liming & Unhair	R		80-116	118-205	1130-2100	140-480	20-50	
	$\bar{X}$		99	148	1680	357	37	>250
	SD		18	50	500	188	15	0.0
Pickling	R			15-90	1300-1310	380-380	32-62	80-206
	$\bar{X}$			52.5	1300	380	47	143
	SD			5.3	70	0.0	21	89
Chrome-Tanning	R	560-1400	98-175	14-30	950-1820	240-510	26-51	46-200
	$\bar{X}$	919	128	197	1357	350	78	128
	SD	372	41.4	896	440	114	82	109
Final Waste	R	50-141	98-120	25-70	.8-1.3	100-300	24-35	68-100
	$\bar{X}$	97	110	53.7	1010	206	29.8	58
	SD	37.4	11	25	200	98.8	4.5	32

$\bar{X}$ Mean of four observations R= Range SD= Standard Deviation

is 180 cm tall with a cross-sectional area of 400 cm^2. The filters are provided with a perforated tray at the top to permit even flow distribution. The plastic media used in the study are made of polypropylene (filter pack, Mass Transfer, Kendal, England). The physical characteristics of the media are: specific surface area 118 m^3/m^2, volume void ratio 0.93 and minimum irrigation rate 5.3 m^3/m^2d. The media are packed randomly in the filters. The activated sludge unit used in the study is described elsewhere[7].

A schematic of the experimental phases of the study is illustrated in Figure 3, Preliminary screening involved removal of particulate matter, using a 1.5 mm mesh screen. Following plain sedimentation for 12 hours, the supernatant was used in a series of jar tests to determine the optimum dose of coagulants and pH level. The supernatant from the coagulation/sedimentation unit was fed continuously to both the biofilters and the activated sludge unit. The effluents of the biological treatment units were further treated in a double-stage filtration system. The filters comprise plexiglass columns 12 cm in diameter and 185 cm in length. The first filter contains sand with 0.2 mm effective size and 6.5 uniformity coefficient.

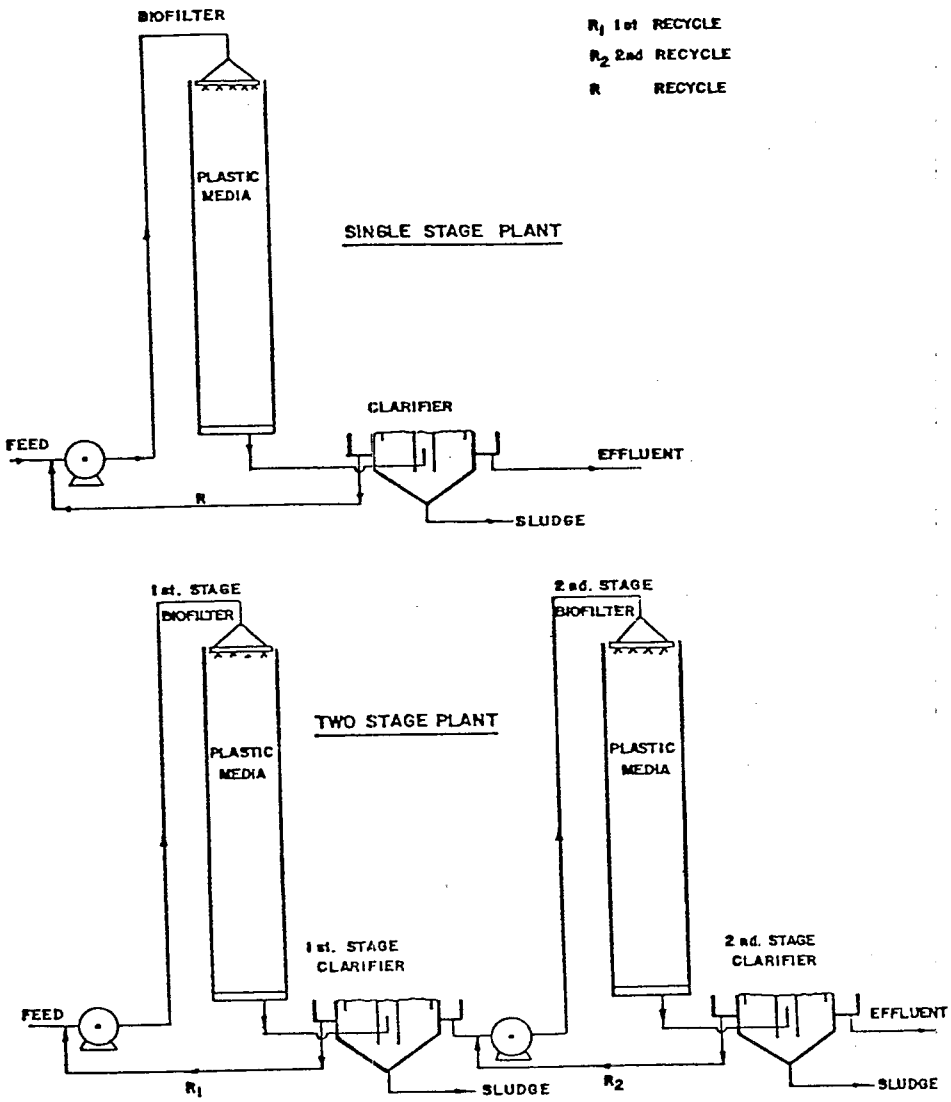


Figure 2. Schematic of Single-Stage and Double-Stage Biofilters.

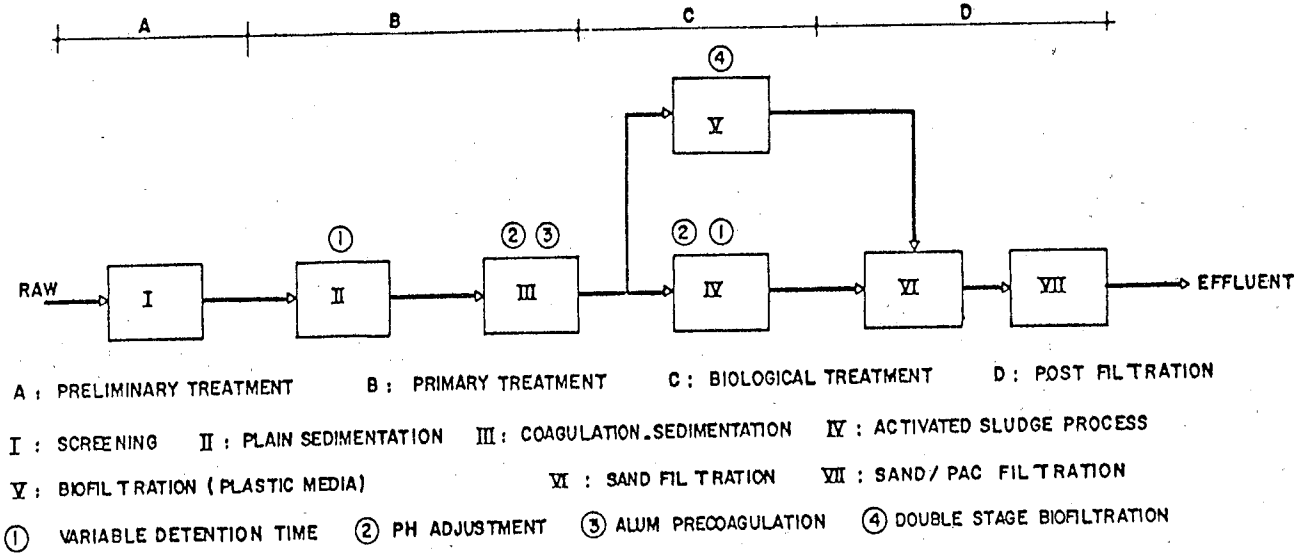


Figure 3. Schmatic of Process Trains for Treatment of Tannery Wastewater.

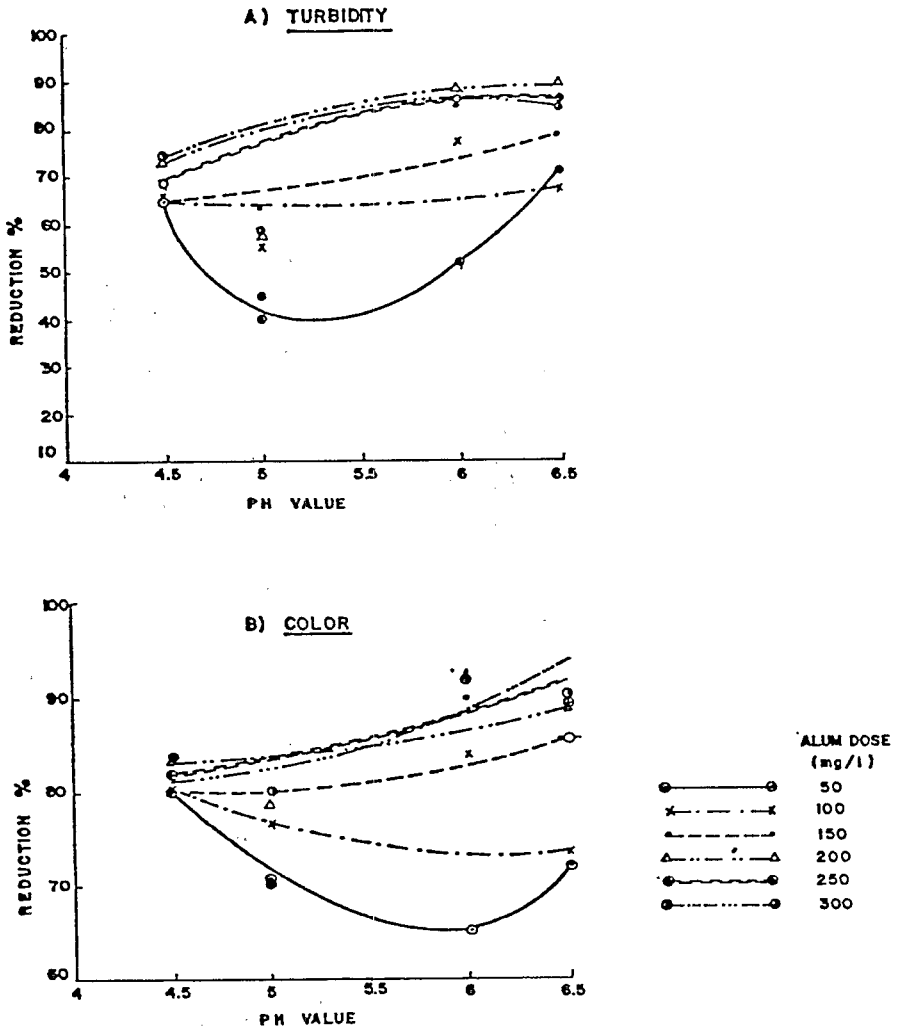


Figure 4. Effect of pH and Alum Dose on the Removal of Turbidity and Color of Tannery Wastes.

The second filter is packed with the same sand mixed with 1% w/w powdered activated carbon.

The physico-chemical characteristics and trace metal analyses of the raw and treated wastewaters were determined according to the procedures described in the Standard Methods[8].

RESULTS AND DISCUSSION

Treatability of Tannery Wastewater

An initial survey indicated that iron salts are not suitable coagulants, due to formation of intense black color. It is presumed that iron salts react with gallic acid in the tannery wastes to form this persistent color. Formation of a turbid, muddy-looking solution and non-settleable sludge precluded the use of lime for pre-coagulation of tannery wastewaters. As shown in Figure 4, appreciable removals of color and turbidity were achieved by alum ($Al_2(SO_4)_3$, 18 H_2O) in the range of 200-300 mg/l at a pH range of 6-6.5. Alum coagulation was also effective for removal of organic constituents and trace metals (Table V).

Recycling of heavily tannery wastewater during biofiltration is indispensable as it appreciably improves the treatment performance. Figure 5 illustrates BOD removal, where recycling

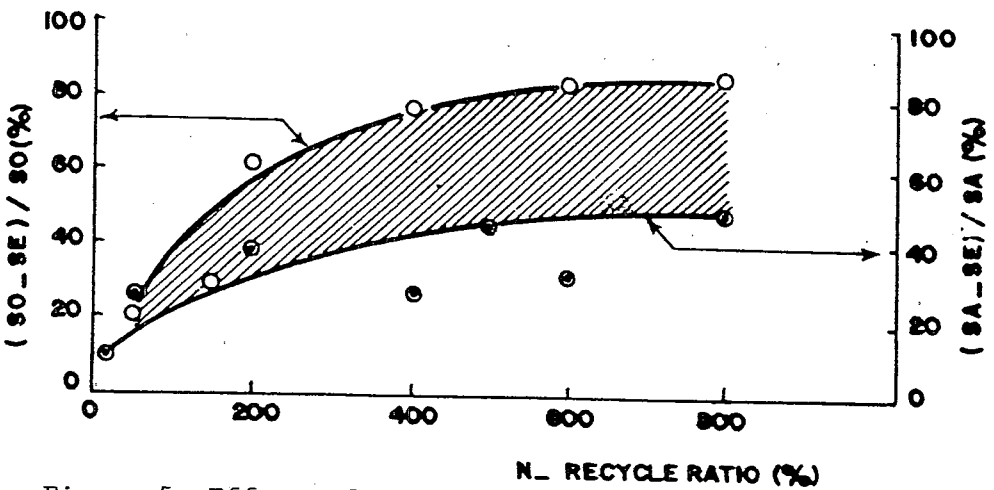


Figure 5. Effect of Recycle Ratio on Apparent and True BOD Removal by Biofiltration.

Table V : Effect of Alum Coagulation on the Removal of Pollutants from Tanning Wastewater

Item		BOD mg/l	COD mg/l	TOC mg/l	SR mg/l	TKN mg/l	H_2S mg/l	Tannin mg/l	O&G mg/l	T-Cr. mg/l	Trace elements (ug/l) Pb	Cu	Fe	Ni	Cd	Zn
Influent	$\bar{X}$	3453	4595	994	1956	120	143	390	130	15.7	79	51	481	133	24	143
	SD	1520	1911	470	1649	59	186	195	56	11.7	37	36	277	46	15	81
Recovery	%	27.3	26.1	28.3	61.7	25.8	52.4	41.5	52.3	81.5	31.6	49.0	35.1	15.8	16.7	28.7

Mean of 6 observations Alum doses 200-300 mg/l pH range (6-6.5)

Table VI :Biofiltration of Tannery Wastewater (mg/l)

A. Single-Stage

Item			TR	VR	SR	BOD	COD	TOC	NH_3-N	TKN	Tannin	Phenol	O&G	H_2S	T-Cr
Influent	$\bar{X}$		22966	5392	245	3971	4933	992	109.8	216	283	2.4	219	298	1.6
	SD		9551	5961	279	2693	2393	936	116	204	51.3	3.1	278	360	1.8
Recovery	%	(E_1)	13.3	N	46.1	71.3	17.1	69.0	N	29.2	62.9	99.2	85.4	99.8	67.5

B. Double-Stage

Influent	$\bar{X}$		17113	1858	1299	2801	3479	769	138.6	163	217	0.7	90	241	12.1
	SD		4415	624	1009	1416	1024	278	173	163	121	0.3	78	375	13.3
Recovery	%	(E_o)	3.9	40.7	78.6	85.5	47.0	79.7	31.2	39.4	75.2	100	85.4	99.9	94.2
Recovery	%	(E_2)		40.7	60.3	49.5	36.1	34.5	31.2	14.4	33.2	100	0.0	95	82.2

N = No recovery $\bar{X}$ = mean of six observations SD = Standard deviation

E_o = Overall recovery percent of the double - Stage system

E_1^o = Recovey percent of the first-stage only

E_2^1 = Recovery percent of the second-stage only

$$E_o = 100 \left[1 - \left[\frac{1-E_1}{100} \right] \left[1 - \frac{E_2}{100} \right] \right]$$

increases both true removal (So-Se) and apparent removal (So-Se), where So, Se and Sa are the influent, effluent and applied BOD after mixing with recycled flow, respectively. An optimum recycling ratio of 600% was selected for the biofiltration study. Recycling of the precoagulated tannery wastewaters in the single-stage biofilter produced moderate recovery of BOD, TOC, tannin, phenol and trace metals (Table VI). However, the recovery of TKN was comparatively low. The effluent of the single-stage biofilter constitutes high levels of organics and chromium which exceed the emission limitations for discharge into public sewers and the sea. Improved recoveries of various pollutants were achieved in the double-stage system as shown in Table VI and Figure 6. The low recovery of TKN during biofiltration is attributed to the presence of a high concentration of nitrifying NH_3-N in the influent (100-136 mg/l) which is toxic to bacteria and hence retards the nitrification process. The average hydraulic rate in the biofilters was 0.02 l/m^2 which is much lower than the adequate rate for wetting (0.062 l/m^2 .S) as recommended by the manufacturer . The high organic loading and the low hydraulic rate contributed to the observed low recoveries of the biofilters even when using the double-stage system. Application of a higher hydraulic rate in larger installations is expected to enhance wetting and consequently improve the overall treatment efficiency.

To compare the performance of biofiltration with other treatment processes, a concurrent study using the Complete-Mix Activated Sludge (CMAS) system was performed. Table VII shows the results of treatment of tannery effluent by the CMAS system using 24 and 48 hours detention periods. The CMAS operated at BOD loadings of 2.5-4.9 $kg/m^3.d$, while maintaining an average Mixed Liquor Suspended Solids (MLSS) of 2300 mg/l and Sludge Volume Index (SVI) of 67 mg/l.

Aeration for 24 hours resulted in moderate recovery of BOD and TOC; high recovery of tannin, chromium and H_2S; while the TKN removal was low. Extended aeration for 48 hours produced a slight improvement in removal of most pollutants. Doubling the aeration time will result in significant increases in capital and operating costs which are not justified by the minor improvement in treatment efficiency. Figure 7 illustrates the comparative effects of the biofiltration and the CMAS processes. With the exception of COD, both processes produced more or less similar recoveries of pollutants associated with tannery wastewater.

An approach to the evaluation of removal of soluble organics from industrial effluents based on molecular size

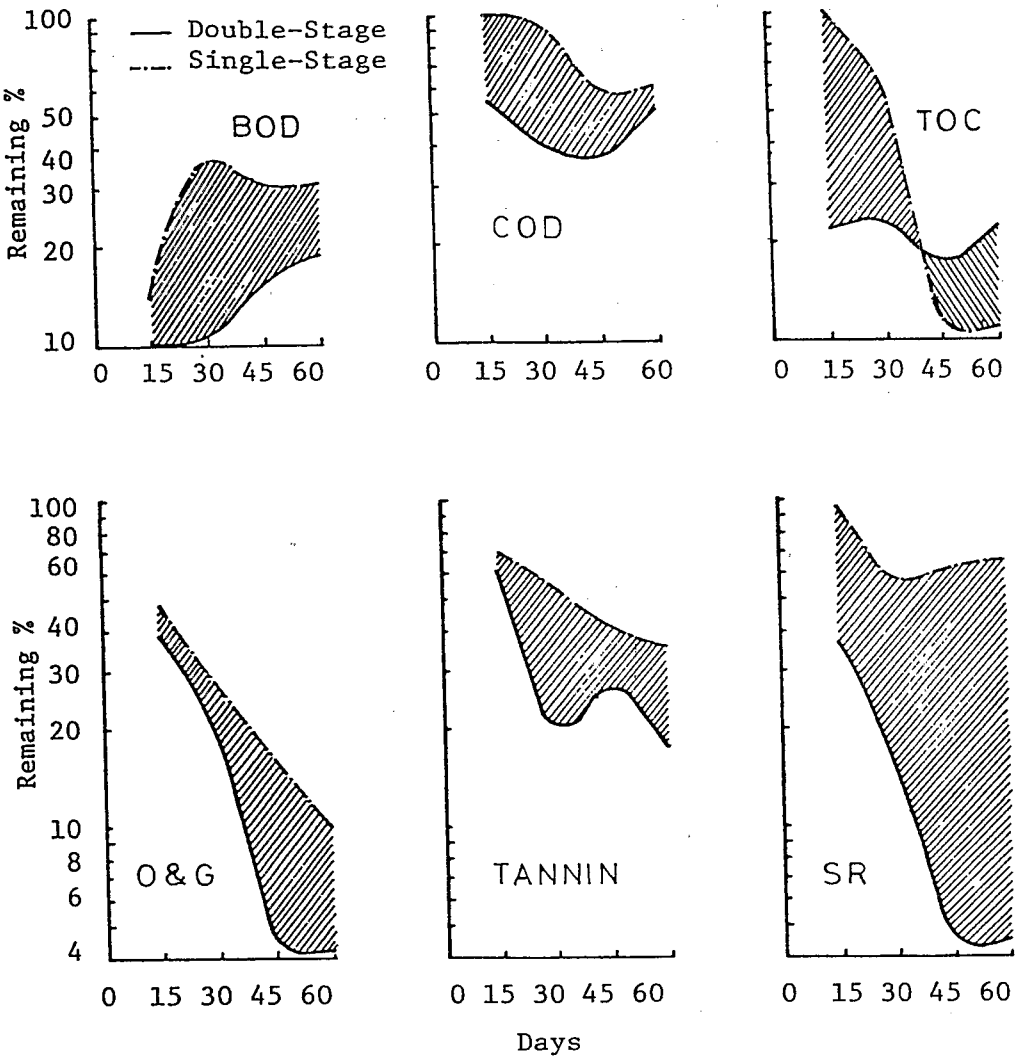
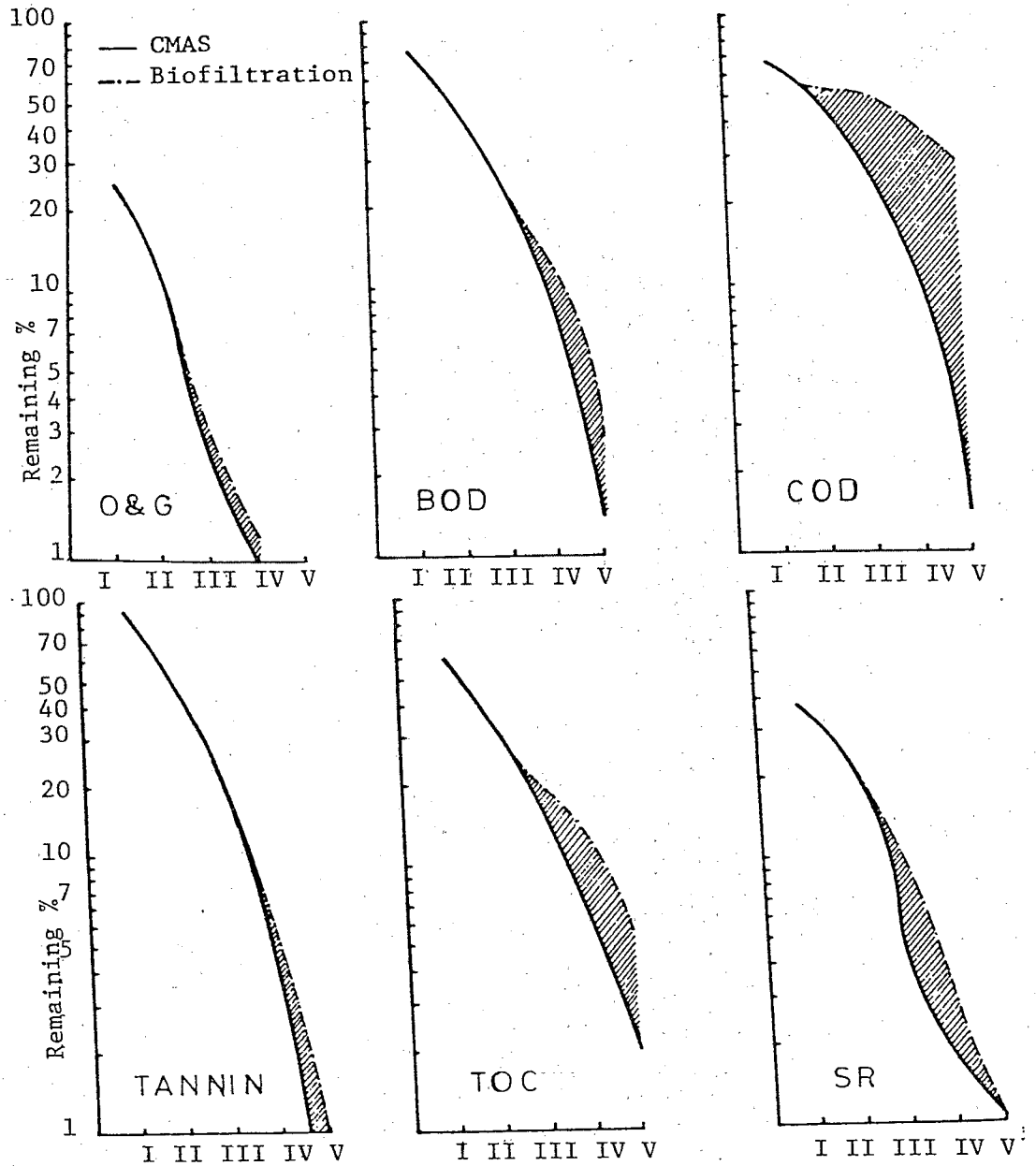


Figure 6. Choronological Effect of Single-Stage and Double-Stage Biofiltration on Removal of Pollutants From Tannery Wastewater.



I. Plain Sed. , II. Precoagulation III. Biological Treatment
IV. Sand Filtration, V. Multi-media Filtration

Figure 7. Effect of CMAS and Biofiltration on Removal of Pollutants of Tannery Wastewater.

Table VII : Effect of CMAS Detention Time on Removal of Pollutents from Tannery Wastewater

Parameter		Influent 24 hr.	Influent 48 hr.	Recovery % 24 hr. (A)	Recovery % 24 hr. (B)	Recovery % 48 hr. (A)	Recovery % 48 hr. (B)
BOD	(mg/l)	2970	2050	70.5	83.0	89.7	92.0
COD	(mg/l)	3166	4028	50.0	82.5	50.9	75.3
TOC	(mg/l)	383	758	63.4	76.0	79.2	90.7
SR	(mg/l)	407	1276	45.5	93.0	80.9	95.5
NH_3-N	(mg/l)	132	152	24.2	34.3	57.2	60.0
TKN	(mg/l)	137	163	11.7	40.0	27.6	55.0
H_2S	(mg/l)	98	5.0	99.7	99.95	100	100
Tannin	(mg/l)	243	213	75.3	86.0	80.3	89.0
Phenol	(mg/l)	0.65	0.33	96.9	98.5	97.6	99.0
O&G	(mg/l)	65	60	80.8	97.6	83.3	98.4
T-Cr.	(mg/l)	3.25	2.0	88.9	99.0	90.0	99.6
Pb	(ug/l)	46	103	16.3	40.3	29.6	69.7
Cu	(ug/l)	24.3	29	17.7	72.5	62.1	87.0
Fe	(ug/l)	340	270	24.3	75.0	52.8	77.3
Ni	(ug/l)	98	173.3	15.3	25.6	37.4	59.0
Cd	(ug/l)	10	20	0.0	31.0	48.5	66.0
Zn	(ug/l)		74.3		61.3	26.0	73.6

(A) Recovery of the CMAS Process only.
(B) Overall Treatment Recovery.

Table VIII :ESTIMATED TREATMENT COSTS OF TANNING WASTEWATER (Costs in US dollars)

PARAMET	ACTIVATED SLUDGE		BIOFJLTRATION	
1. Capital Costs				
a. Primary (Screening, Sed. and Precoagulation).	285,000		285,000	
b. Secondary (biological).	345,000		265,000	
c. Tertiary(mixed media filtration).	205,000		205,000	
d. Non-Component cost (piping, Instruments).	154,000		120,000	
	989,000		875,000	
Amortizations of Capital Costs (20 years)		98,900		87,500
2. Operating Costs				
a. Chemicals	120,000		120,000	
b. Power	22,500		2,500	
c. Labor	15,000		6,000	
d. Maintenance	18,000		5,400	
		175,600		133,900
		274,500		221,400
Percentage costs compared to activated sludge.		100		80.7

distribution using Gel Chromatographic (GC) technique has been detailed by Hamza and Tambo[9]. The organic constituents which are easily eluted by water (Group I) are amenable to biological treatment, while organics with high affinity with the gel (Group II) require elution by NH_4OH. This portion can be effectively removed by tertiary treatment. Schematic GC patterns of raw wastewater and the effluent of the CMAS and biofiltration are illustrated in Figure 8. The GC patterns indicate the presence of relatively high Group II constituents which require further treatment for complete treatment. Mixed media filtration removed 80-85% of the residual Group II constituents.

Treatment Costs

The cost estimates given in Table VIII reflect costs applicable to centralized treatment of tannery effluent to produce water suitable to discharge into the sea. The estimates are based on the projected discharge of 2.5 million cubic meters annually after implementation of the renovation and expansion plan of GOFI. Cost estimates assume 20 years' service life and 5% low interest rate provided by the Egyptian Govovernment for public sector industries. Operating costs were based on prices of 1982, and cost analyses were performed according to EPA Guidelines[10]. As shown in Table 9 the estimated cost of biofiltration is less than that of the CMAS process. Further savings are expected if the clean water pool can be segregated and discharged without treatment.

In case of lack of sufficient financial support, it is proposed to install pre-treatment and biofilters first, while adding the post-filters at later stage.

Toxicity of Raw and Treated Tannery Effluents

The effect of raw wastewater on the survival of Cyprimus carpio and Mugil capito has been investigated as a complementary part of the study[11]. The mean survival time was 5 minutes for hatched embryo, 1.4 hour for larval stage and 17.2 hours for juvenile stage of Cyprimus carpio and 16.4 hours for juvenile stage of Mugil capito. The high toxicity is attributed to the presence of high concentrations of trace metals, organic constituents, H_2S salts and ammonia. The 96 hours LC_{50} of the hatched embryo and larval and juvenile stages of Cyprimus carpio exposed to the biofiltration effluent was achieved in treated effluent diluted with water to 3.1%, 5.2% and 8.3%, respectively. Although the treated waste was less toxic than the raw waste, it is estimated that treated tannery waste must be

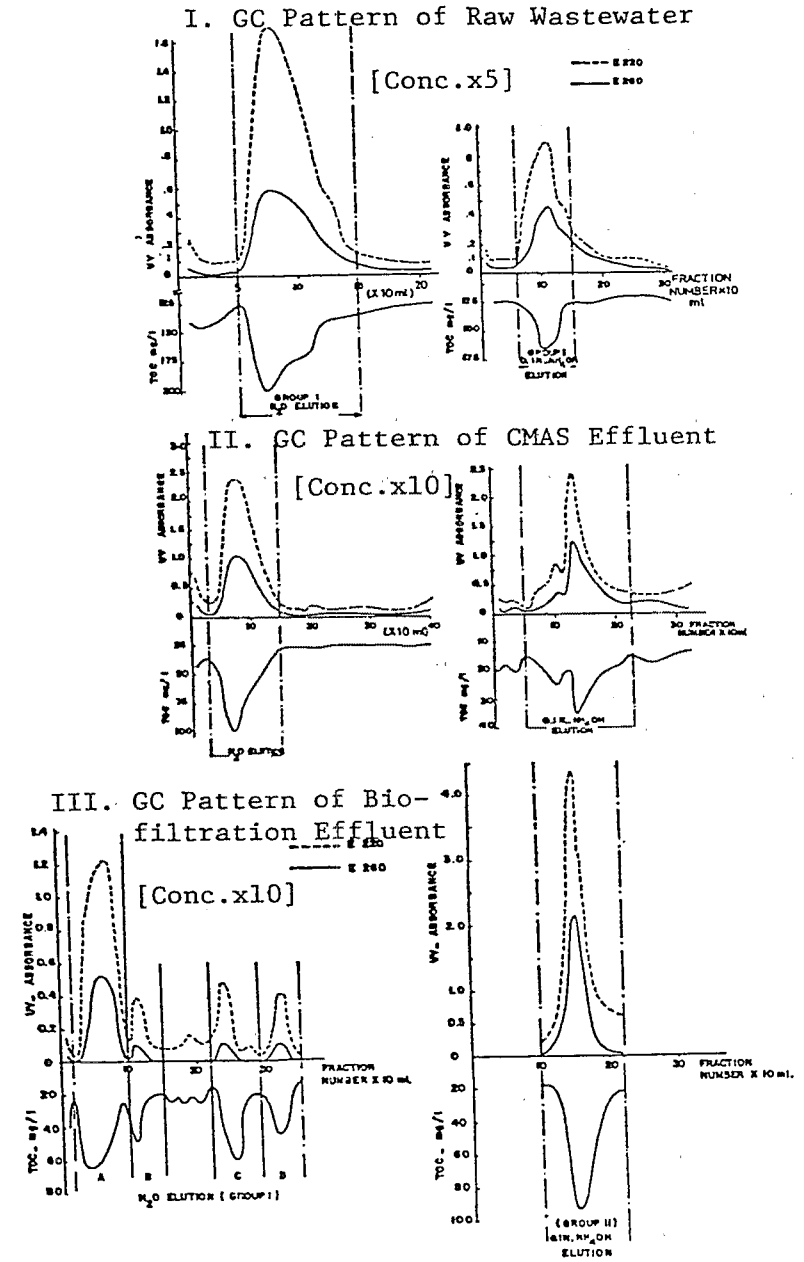


Figure 8. GC Patterns of Raw and Biologically Treated Tannery Effluents.

diluted 40 times to reduce the fish toxicity to an acceptable level. The toxicity of the treated effluent is attributed to the high concentration of NH_3 (77 mg/l) which causes severe histopathological changes in gill structure. Ammonia stripping is being investigated as an option to reduce fish toxicity.

CONCLUSIONS

A centralized treatment system is proposed for the MIC tanneries. The treatment train encompasses precoagulations, biofiltration and post-filtration. Among the treatment alternatives, biofiltration is the most adequate due to lower operating costs, suitability for intermittent flow, versatility and relative ease of operation. The proposed treatment train complies with the limitations for emission into the sea (except for NH_3). Reduction of the toxicity of the treated effluent can be achieved by ammonia stripping or by dilution when mixed with sea water. Segregation of non-polluted effluents, originated in clean water pools, from polluted wastes is expected to reduce treatment costs. Government subsidization of the centralized treatment facility is necessary to encourage MIC tanneries to institute the proposed treatment system.

ACKNOWLEDGEMENTS

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PART X: INNOVATIVE RESEARCH

Effect of Periodic Flow Reversal Upon RBC Performance

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Introduction

Rotating Biological Contactors are traditionally operated as a series of units through which water always flows the same direction. After the microbial film is established on the RBC's, a characteristic pattern of growth is seen across the stages: growth is heaviest on the first stages and diminishes with each sucessive stage once the organic concentration falls to the level below which removal is a function of organic concentration. The biological communities on each stage change in response to their differing environments. The total growth on a stage is roughly proportional to the organic concentrations which normally occur within it.

When shock loadings occur across multistage RBC units, downstream stages receive higher organic loadings than usual. Studies of RBC response to shock loadings have revealed that downstream stages have only a limited capacity to treat these short duration excess loadings. Much of a shock load passes through an RBC untreated.

The purpose of this research was to explore the feasibility of increasing an RBC's capacity to treat shock loads by periodically reversing the direction of flow across two or more successive stages of an RBC installation. This concept was first suggested by Borchardt, et al., in June of 1978 as a result of their studies of RBC nitrification.[1] RBC biofilms have been observed to grow more rapidly in response to an incremental organic concentration than they decay in response to an equal decrement. This characteristic of biofilms suggests the possibility of increasing the total inventory of film within several stages of an RBC by periodically reversing the direction of flow across those stages. The period of reversal would be sufficiently long to permit the concentration gradient across the stages to re-establish itself in each direction but not so long that the biofilm could adjust fully through growth and decay to the new distribution of nutrients. Under average flow conditions, the RBC's performance shouldn't be significantly affected. The former earliest stage would see a lower average concentration than under conventional flow and would presumably perform less removal. However, the former latter stages would see higher average concentrations and due to the advantage of growth over decay they would develop heavier growth which would remove those organics now passed by the first stage. Far more reserve capacity would exist to treat shock loads due to the greater inventory and more even distribution of biofilm created by periodic reversal.

An analogous mode of operation was explored for trickling filters in England forty years ago.[2] Flow was reversed periodically across two filters in series in order to prevent ponding on the more heavily loaded filter. The alternate heavy feeding and comparative starvation experienced by the slime encouraged first growth and then endogenous respiration and sloughing. Under the "alternating double filtration" mode of operation, much less buildup of slime and consequent ponding was observed. When slowly rotating distributer arms came into vogue, alternate feeding and starvation became feasible within one filter. Ponding was seldom a problem and the alternating double filtration mode of operation died out. Its implications for treatment efficiency and effluent variability were never explored.

This research will evaluate the flow reversal concept in RBC's. Today we will report our preliminary results. More elaborate follow-up experimentation is now underway.

Materials and Methods

A Clow pilot scale RBC was used in this research. The 13 foot (4 m) diameter disks had a total area of 11000 sq. ft. (1022 m^2). The media was set in a 2000 gallon (7.6 m^3) tank divided into four equal compartments to allow staging. The media rotated at about 1.6 rpm.

For purposes of this experiment, the plant was configured as two 2-stage RBC's in parallel. Only one pair was actually sampled.

We assumed that for flow reversal to have a significant effect, organic concentrations in the reversed stages would have to be low enough so that removals achieved depend on their fluctuations. In some plants, the first stage or two is saturated with respect to organics. Higher organic concentrations do not produce appreciably higher removals on a mass basis. Percent removals fall. At these high organic concentrations, the mass transport of oxygen into or waste products out of the film is controlling. The source of wastewater for this experimentation was chosen to be the partially treated effluent of an Imhoff tank. This moderate strength wastewater (BOD_5 of approximately 100 mg/l) never saturated the first stage of the Clow Corporation pilot scale RBC plant used for the reversal experiments.

Wastewater was pumped through the pilot RBC at flow rates between 28,800 and 72,000 gpd (272 m^3/ day). The flow was measured with an in-pipe flow sensor. Hydraulic loadings varied from 2.6 to 6.5 gpd/ft ^{2} (.107 to .267 m^3/m^2/day). The corresponding organic loadings were 1.67 to 4.17 lb SCOD/1000 ft^2/day. The soluble ultimate BOD to soluble COD ratio was variable but averaged .6.

The influent and effluent wastestreams were sampled every hour with ISCO automatic samplers. Two ml of sulfuric acid were added to each sample bottle before sample collection to stabilize the sample. No sample was more than 26 hours old when it was analyzed.

The Hach Reactor Digestion COD method was used to assess the RBC's performance. This is an EPA approved method which is convenient for the rapid analysis of large numbers of samples. All samples were filtered through Whatman No. 5 before analysis. ThepH and temperature were measured daily while experimentation was underway. Dissolved oxygen rose from about 4 mg/l in the influent to near saturation in the effluent at the lowest organic loadings. The dissolved oxygen profile across the system was essentially flat at the highest loadings.

Experimentation was performed in three phases. First the RBC was operated in the conventional manner for a period of almost two months. The flow rate during this period of film establishment and maturation and base-line data collection was 28,800 gpd (109 m^3/day). Good removals (about 60 percent of the influent COD) were obtained by three weeks after startup. A severe thunderstorm then scoured off much of the growth requiring us to install a cover and allow some regrowth prior to experimental data collection.

The second phase of the experiment involved daily flow reversal. During this phase, valves were opened and closed immediately after the noon sampling to make the first stage the second, and the second, the first. Flow rates of 28,800 gpd (109 m^3/day), 57600 gpd (218 m^3/day) and 72,000 gpd (273 m^3/day) were employed during this phase to allow the identification of any interactions between hydraulic loading and flow reversal. Reversal experiments continued for seven weeks.

The final phase of experimentation involved a return to conventional one way flow. During this phase, the higher flow rates used for reversed flow studies were also applied with conventional flow.

Results and Discussion

The results of the experiments performed are presented in Figure 1 and summarized in Table 1. Flow reversal was initiated after almost 2 months of consistent COD removals with conventional flow. Percent removals rose steadily during the first 5 weeks of reversal despite greatly increased hydraulic loadings. This was a striking result since

hydraulic loading and percent removal are normally either constant or inversely related in a moderately loaded RBC.[3] When the hydraulic loading was reduced to the pre-reversal level for the purpose of obtaining reversed flow performance data directly comparable to the conventional flow baseline, sloughing increased markedly. Percent removals fell sharply. The earlier higher flow rate was restored and conventional flow operation was resumed when this trend became apparent so that the expected gradual return of the system to baseline performance might be observed. Performance was falling at approximately the same rate as it had earlier increased when a violent thunderstorm blew the unit's canvas cover into the tank and broke the coupling between the driven gear and the RBC shaft. Before repairs could be made and a suitable replacement cover procured, the unit froze up for the winter, prematurely terminating experimentation.

Table 1

Results of Experimentation

		Fraction Removed	
Type Data	# Observations	$\bar{x}$	sd.
one way, 20 gpm	64	.473	.132
reversal, 20 gpm	55	.575	.149
reversal, 40 gpm	40	.625	.086
reversal, 50 gpm	71	.636	.045
reversal, 40 gpm	54	.722	.053
reversal, 20 gpm	42	.488	.095
one way, 40 gpm	36	.641	.082
one way, 50 gpm	72	.574	.073

During these preliminary experiments, an average of 55 percent of the applied COD was removed under conventional flow operation and 62 percent was removed with reversal. The difference in the means was significant at the 1 percent level based on 436 influent/effluent sample pairs. Influent concentrations showed no correlation with percent removals.

The reversed flow effluents were less variable than were the conventional flow effluents, especially when the

initial conventional flow baseline data were compared to the reversed flow data. However, the experiments were terminated before the higher flow rate conventional flow effluents reached a steady state. Therefore, these data do not establish a difference in effluent variability.

Experiments are underway now in which two compartments of the Clow pilot plant are operated as a conventional two-stage RBC and the remaining two are reversed daily. The conventional flow control and the reversal experiment will receive an identical influent and will operate under identical climatic conditions for several months. An unequivocal comparison of the two modes of operation will then be possible. If the benefits of flow reversal suggested by the early research are confirmed by this follow-up work, optimal reversal periods can be identified and economic feasibility analyses can be performed.

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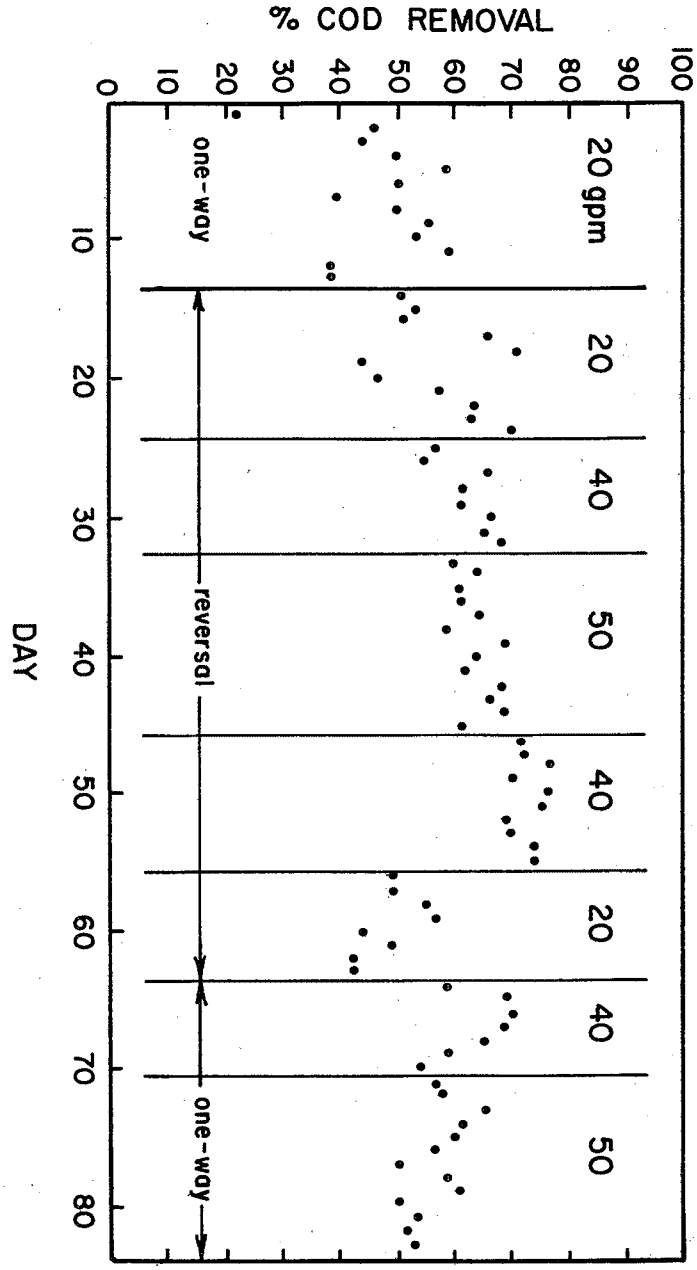

Figure 1. Experimental Results (Daily Averages).

AN ASSESSMENT OF DISSOLVED OXYGEN LIMITATIONS AND INTERSTAGE DESIGN IN ROTATING BIOLOGICAL CONTACTOR (RBC) SYSTEMS

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INTRODUCTION

Over 260 RBC systems are presently in operation in this country, with flowrates ranging from less than 0.1 to 54MGD (1). RBC systems have received much attention in recent years, including investigations by EPA of process, power, and equipment performance (2), and a National RBC Symposium in 1980 (3).

Numerous reports in the literature and RBC facility surveys (1, 4, 5, 6, 7, 8) have reported difficulties with the initial stages of RBC systems reflected by heavy biofilm growth, the presence of nuisance organisms (beggiotoa), and a reduction in organic removal rates. These problems have been attributed to excessive organic loadings which result in dissolved oxygen deficiencies in the biofilm. Recent unpublished surveys have reported results (9) which indicate that nuisance organism growth may be precipitated exclusively by high sulfide concentrations.

Empirical design approaches proposed by RBC vendors are presently used almost exclusively by engineers for RBC municipal wastewater treatment design. Only one vendor design presently makes reference to the aforementioned organic overload condition (10). A loading limitation is recommended and defined in terms

of a limiting organic loading (e.g. for a mechanical drive RBC system, 4.0 lbs/day/1000 sf of soluble BOD loading on the first stage).

Recent literature (11, 12) have stressed the need for design techniques based upon fundamental principles of substrate and oxygen, transport and utilization to help define oxygen limitations and stagewise kinetic parameters. The development of a mechanistically sound kinetic model for RBC's or any fixed-film biological process, however, is extremely difficult due to the complex relationships among transport and other kinetic phenomenon which must account for both substrate and oxygen transport and utilization. The use of these complex models by design engineers is not practical.

The empirical vendor approaches are simple to use, requiring the input of flowrate and organic concentration to establish a given design surface area. They are limited, however, in that they have no fundamental basis to account for organic removal, oxygen limitations, staging requirements, step feeding or recycle, and do not address interstage removal.

On the basis of a review of available design methodologies, it is apparent that a simplified approach is desireable to define RBC performance on the basis of rational kinetics. The approach described by Clarke et. al. (11) is a reasonable method which defines interstage removal on the basis of Monod growth functions, and a mass balance with respect to substrate across a completely mixed reactor. No simple model, however, can account for the interaction between oxygen requirements and substrate utilization. The design of RBC systems at the present time, then, is best undertaken in a simplified two-step approach:

Step 1. Define limiting design conditions to prevent organic overloading and dissolved oxygen limitation.

Step 2. Utilization of a rational design method to define the substrate removal capacity of the RBC surface area under conditions which are not oxygen limiting.

Under an ongoing contract with the Municipal Environmental Research Laboratory of the U.S. Environmental Protection Agency, data which define organic overload (oxygen limiting) conditions were sought and an interstage model calibrated from field data collected from several operating facilities.

ORGANIC OVERLOAD CONDITIONS[1]

Optimum RBC process design requires that the microbial substrate utilization rate represent the rate limiting condition. Under this condition, the process achieves maximum use of surface area because substrate removal is limited by the ability of the biomass to assimilate the waste and no extraneous factors (e.g. lack of dissolved oxygen) limit the rate of substrate removal.

Except for the aforementioned vendor recommendations concerning organic loading limitations, little data is available to define the conditions which induce dissolved oxygen limitations. Williamson and McCarty (12), utilizing their fixed film model, predicted that dissolved oxygen limitations would occur at soluble BOD (SBOD) concentrations of 40 mg/l, adjacent to the biofilm. This would correspond to a mixed liquor RBC reactor concentration somewhat greater than 40 mg/l SBOD. Field observations, literature reviews, and telephone interviews were made to determine influent conditions which result in organic overloads. Overloaded conditions on an RBC were identified by the characteristic colonization of the media by nuisance organisms which gain a competitive advantage over other organisms under oxygen deficient conditions. For each of the facilities surveyed, information concerning nuisance organisms, influent organic concentrations, and hydraulic loading were recorded to determine wastewater characteristics which could be associated with these conditions. The survey included a total of twenty-three facilities. Results are tabulated in Table I.

A graphic presentation of these results is presented in Figure 1, which depicts a plot of total influent BOD on the ordinate and first-stage hydraulic loading on the abscissa. The graph presents a demarcation line that separates facilities which experienced first-stage overloading problems from those which did not (i.e. by the presence or absence of nuisance organism).

The relationship is depicted as a hyperbolic function, where the product of the variables is a constant:

(BOD)(Hydraulic Loading) = Constant Organic Loading

From the graph, it can be seen that the organic loading that separates plants with overloaded conditions from plants without such problems is 6.4 pounds BOD/day/1000 sf.

---

[1] Organic overload and oxygen limitations are used interchangeably.

TABLE I. Organic Overloading Conditions Related to Influent Organic Concentration and Hydraulic Loading

Plant[1] ID No.	Average RBC Influent BOD Concentration (mg/l)	Average 1st Stage Hydraulic Loading (gpd/sf)	Calculated Average 1st Stage Organic Loading (lbs/day/1000sf)	DO-Limiting[2] Condition
1	125	7.4	7.4	P
2	48	3.6	1.4	A
4	50	4.0	1.7	A
5	182	10.1	15.3	P
8	47	15.5	6.1	A
10	169	4.7	6.6	P
12	85	12.5	8.9	P
13	55	9.1	4.2	P
14	96	5.0	4.0	A
15	98	9.6	7.8	P
16	72	6.6	4.0	A
29	93	18.5	14.3	P
30	175	7.2	10.5	P
31	145	6.2	7.5	P
32	505	1.9	8.0	P
33	350	3.9	11.4	P
34	336	5.9	16.7	P
35	180	6.8	10.2	P
3	118	5.1	5.0	A
6	152	3.9	4.9	A
9	144	7.9	9.5	P
11	96	9.0	7.2	P
36	213	6.5	11.5	P

[1]: Except for air driven plant no. 16, all are mechanical drive facilities with no supplemental air.

[2]A: Plants experiencing no nuisance organism growth (absence) in the First Stage.

[2]P: Plants experiencing problems with nuisance organism growths in the First Stage.

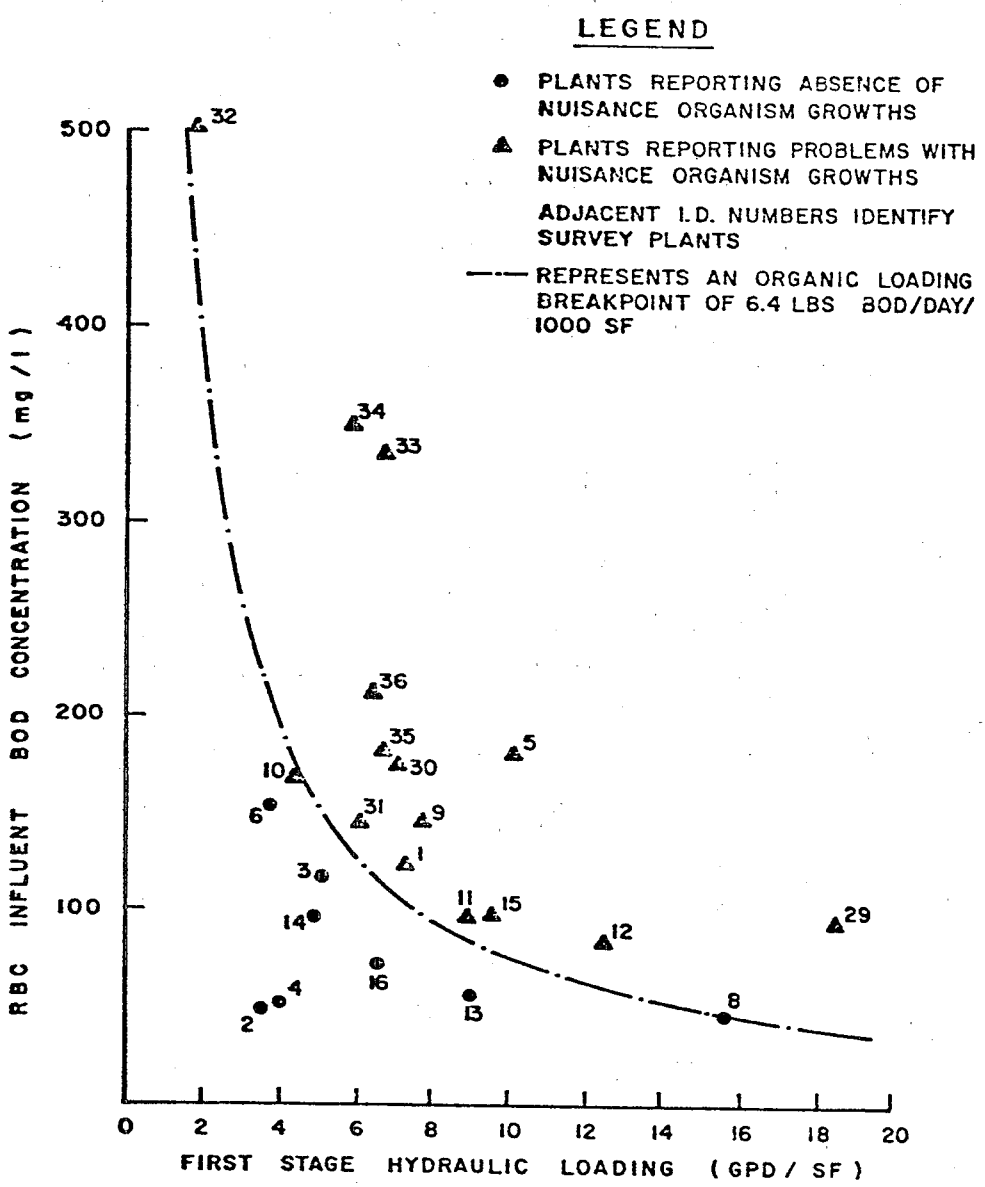

FIGURE 1: D.O. LIMITING CONDITIONS RELATED TO INFLUENT ORGANIC CONCENTRATION AND HYDRAULIC LOADING

INTERSTAGE MODEL CALIBRATION

The Clarke model previously mentioned is based upon a mass balance with respect to substrate across an assumed completely mixed RBC stage at steady-state which can be written as follows:

$$V\left(\frac{ds}{dt}\right) = FS_o - FS_1 - \frac{\mu}{Y} X A \qquad \text{(Equation 1)}$$

where,

V = reactor liquid volume (volume)

$\frac{ds}{dt}$ = change of substrate concentration with time (mass/volume · time)

F = wastewater flow rate (volume/time)

S_o = influent organic concentration (mass/volume)

S_1 = effluent organic concentration (mass/volume)

μ = specific growth rate of attached RBC microorganisms (^{1}/time)

Y = apparent yield of attached RBC microorganisms $\left(\frac{\text{mass biomass produced}}{\text{mass substrate consumed}}\right)$

X = mass of attached microorganisms per unit area (mass/area)

A = RBC surface area (area)

This equation assumes that the intensity of mixing in each stage is sufficient for complete mixing, and that organisms decay is small compared to the growth rate, and that all substrate removal is due to attached biomass.

Using the Monod growth function:

$$\mu = \mu_{max} \left(\frac{S_1}{K_s + S_1}\right) \quad \text{and,} \qquad \text{(Equation 2)}$$

$$V\left(\frac{ds}{dt}\right) = FS_o - FS_1 - \frac{\mu_{max}}{Y} A X \frac{S_1}{K_s + S_1} \qquad \text{(Equation 3)}$$

defining,

$$P = \frac{\mu_{max}}{Y} X,$$ as the area capacity constant (i.e. maximum substrate which could be removed per unit area per unit time) (Equation 4)

and,

$$R = F(S_o - S_1) / A,$$ as the removal coeffiecient (actual substrate removed per unit area per unit time) (Equation 5)

then,

at steady-state $V(\frac{ds}{dt}) = 0$

and,

$$R = \frac{PS_1}{K_s + S_1}, \text{ or} \quad \text{(Equation 6)}$$

$$\frac{1}{R} = \frac{K_s}{P} (\frac{1}{S_1}) + \frac{1}{P} \quad \text{(Equation 7)}$$

A major feature of the Clarke model is its use of a rational approach for defining substrate removal converting the terminology into design parameters which are readily used in the field today: R, the removal coefficient which reflects the organic removal rate, can be defined in terms of pounds/day/1000 sf, and P, the area capacity constant, which represents the maximum removal rate possible, can also be described in terms of pounds/day/1000 sf.

To calibrate the Clarke model, interstage soluble BOD[1] data was collected from eleven RBC facilities selected with three to six RBC stages where organic overload (i.e. greater than 6.4 lbs/day/1000 sf BOD organic loading) did not exist. Organically overloaded facilities were screened for two reasons:

1. the organic removal rate of these facilities are highly variable as a result of nuisance organism interference and the influence of DO deficiencies on substrate removal rate; and,
2. the intent of the designer as presented is to avoid this condition (i.e. insure that design is not organically overloaded by keeping the loading level below the DO-limiting level).

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[1] Interstage data available at the eleven facilities was soluble BOD (SBOD).

The eleven facilities provided influent values of soluble BOD from 10 - 95 mg/l, with a 55 mg/l average value. Hydraulic loadings ranged from 0.4 - 1.5 gpd/sf with a 1.3 gpd/sf average value.

Using both the organic removal per unit area, R, and the soluble BOD concentration in the reactor, S, the model was calibrated as follows:

> Values of $(^1/R)$ vs $(^1/S)$ for each stage of all eleven plants were plotted to yield a straight line with a slope of K_s/P and a y-intercept of $^1/P$, per Equation 7.
>
> The maximum removal rate, P, and the half velocity coefficient, K_s, were computed from this graph.

This graphical analysis for the four consecutive stages is presented in Figures 2 and 3, respectively.[2] The results of the four consecutive stages are presented in Table 2.

TABLE 2. Calibrated Maximum Removal Rate, P, and Half Velocity Coefficients, K_s (Soluble BOD)

Stage No.	Maximum Removal Rate, P		Half Velocity Coefficient, K_s
	(GPD/SF · mg/l)	(Lb/Day/1000 SF)	(mg/l)
1	1000	8.34	161
2	667	5.56	139
3	400	3.34	82
4	100	0.33	25

[2]Straight lines were drawn through the data, visually weighting the distribution of data points and screening data which were judged as outliers.

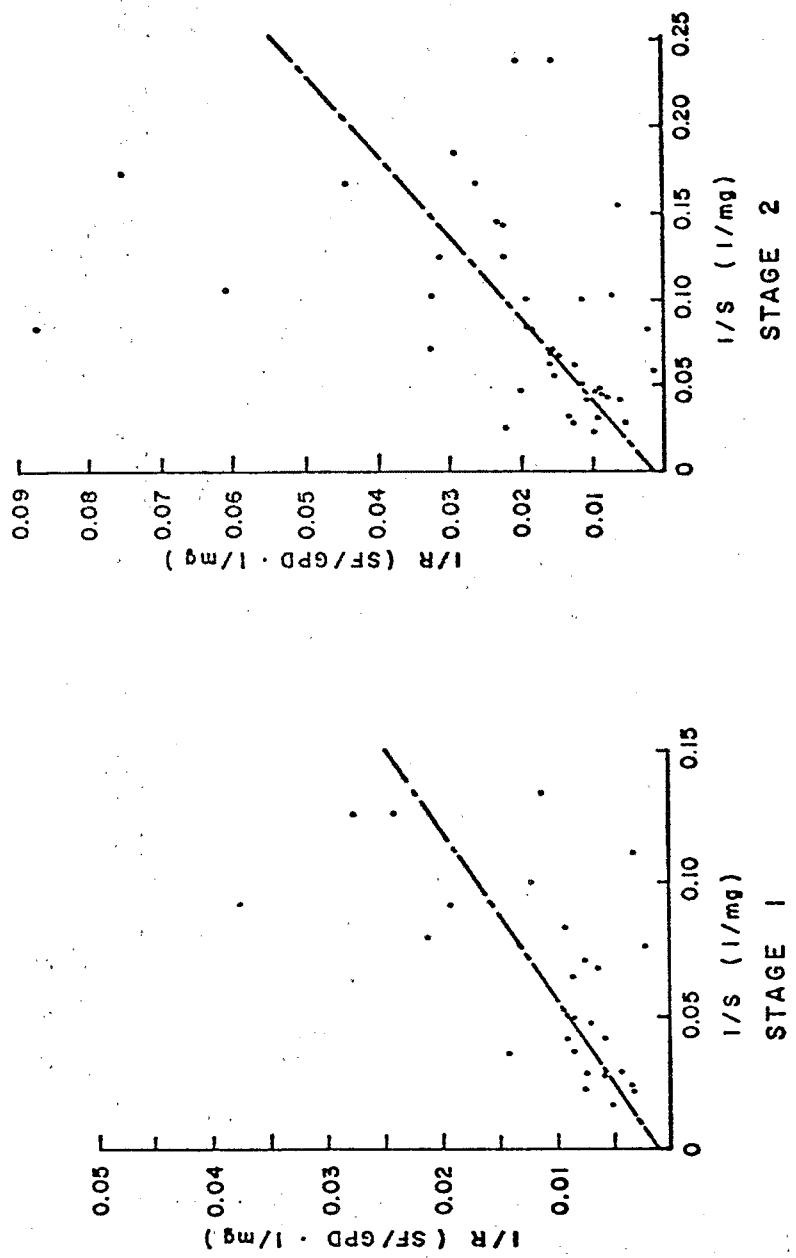

FIGURE 2: DATA ANALYSIS, I/R VS. I/S

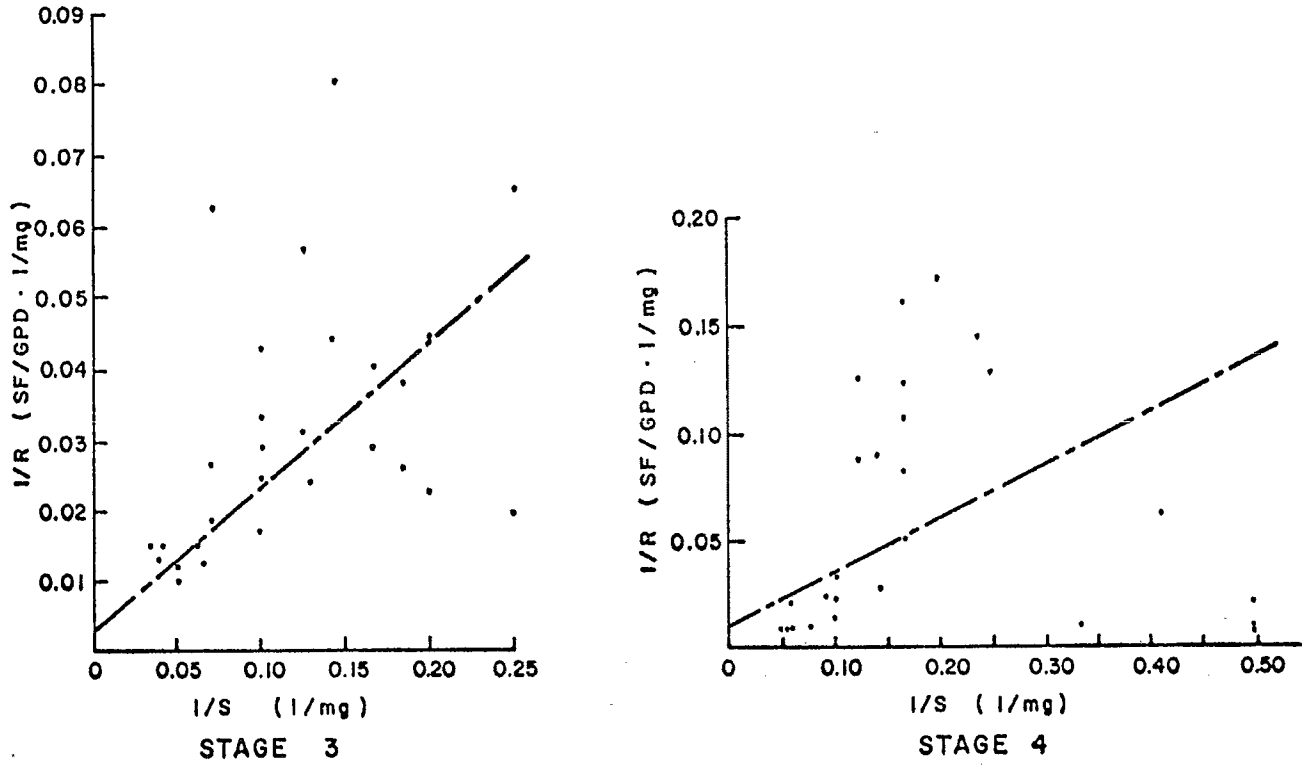

FIGURE 3: DATA ANALYSIS, I/R VS. I/S

Values of effluent concentration in each stage can be expressed as a function of flowrate, influent concentration and surface area by setting Equation 5 equal to Equation 6 and solving for the effluent concentration, S_i, as follows:

$$R = \frac{F_i}{A_i}(S_o - S_i) = \frac{P_i S_i}{K_s + S_i} \qquad \text{(Equation 8)}$$

The effluent concentration for any stage can be expressed as:

$$S_i = \frac{[[HL_i(S_o - K_i)] - P_i] + \sqrt{[HL_i(S_o - K_i) - P_i]^2 + [4(HL_i)^2](K_i \times S_o)}}{2(HL_i)} \qquad \text{(Equation 9)}$$

where,

S_o = influent SBOD concentration (mg/l)

HL_i = hydraulic loading (gpd/sf)

S_i = effluent SBOD concentration (mg/l)

P_i = area capacity constant (1/day)

K_i = half velocity coefficient (mg/l)

F_i = wastewater flowrate (gpd)

A_i = area (square feet)

i = denotes RBC stage under analysis

To determine the hydraulic loading given the influent concentration and the required effluent concentration, Equation 8 can be rearranged as follows:

$$HL_i = \frac{(P_i)(S_i)}{(K_i + S_i)(S_o - S_i)} \qquad \text{(Equation 10)}$$

To assess the accuracy of the calibrated interstage (4 stage) model, a comparison between the observed field performance of sixteen RBC facilities and the model was undertaken.[1]

The sixteen facilities were divided into influent concentration ranges of 52, 97, 148 and 214 mg/l.[2] In order to compare the model, which expresses BOD in terms of soluble BOD (SBOD), an SBOD:BOD ratio of 0.5 was assumed

The model is compared to regression lines of field performance in Figure 4. The top graphic of Figure 4 presents the data as organic removal in lbs/day/1000 sf vs hydraulic loading in gpd/sf; the bottom graphic as organic removal vs organic loading, both in lbs/day/1000 sf. The regression line extends across the range of loading conditions (hydraulic and organic) observed in the field. From Figure 4, it can be seen that the model is more accurate (i.e. with respect to the regression line) for the lower concentration ranges, the lower hydraulic loading ranges and the lower organic loading ranges.

The lack of correlation at higher concentrations and loadings can be attributed to the fact that:

a. the model uses the Monod growth function to account for both substrate utilization and mass transfer. In fixed film systems mass transfer, however, is dependent upon reactor concentration (i.e. actually the concentration adjacent to the film). Higher influent concentrations would exhibit greater mass transfer driving forces which would increase values of P and K_s calibrated in a substrate deficient system.

b. the model was not designed to predict removal in organic overloaded environments and at the higher loadings where the model does not correlate well, first stage loadings may exceed design levels (e.g. greater than 6.4 lbs/day/1000 sf).

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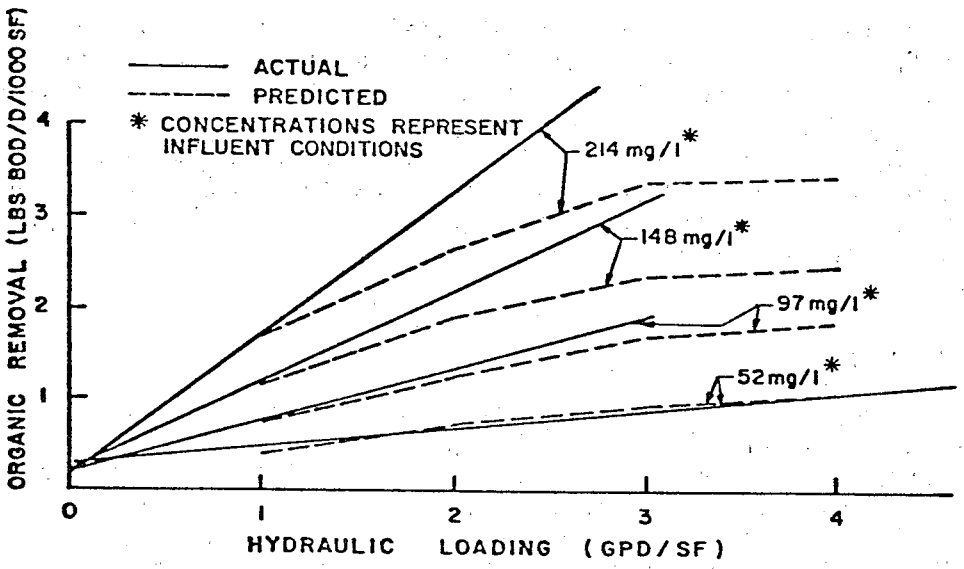
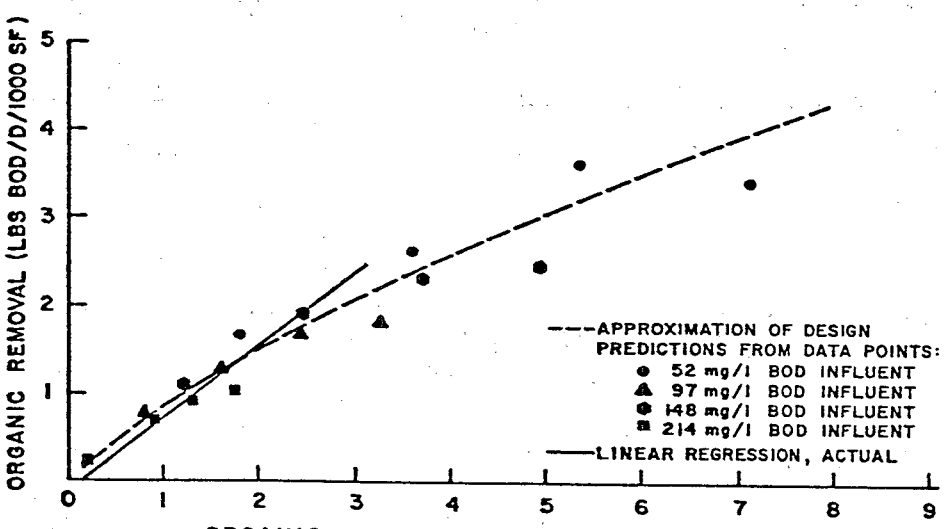
[1] Primary clarifier BOD removal was assumed to be 30% for each facility with primary clarifiers.

[2] Detailed data concerning these sixteen facilities are presented elsewhere (2).



ORGANIC REMOVAL VS. HYDRAULIC LOADING



ORGANIC REMOVAL VS. ORGANIC LOADING

FIGURE 4: COMPARISON OF ACTUAL PERFORMANCE TO DESIGN PREDICTIONS

MODEL ANALYSIS

The major advantage of a model such as the one described and calibrated here is its ability to predict the effect of changes in system characteristics, such as staging and step feeding, upon RBC performance. In performing these analyses, it must be recalled that the model was intentionally calibrated under conditions which were organically underloaded. As a result, advantages or disadvantages of various process variations apply only to those conditions.

Staging

Assuming an influent BOD = 100 mg/l (i.e. SBOD = 50 mg/l), and a flowrate of 1.0 million gallons per day, the interstage model was used to predict the benefits of staging by assessing the amount of surface area required to achieve efficiencies of 50, 60, 75, 85, and 95 percent. The results illustrated in Figure 5 indicate that multistage systems require less surface area than single stage systems to achieve a given percentage removal, up to a point. For the influent condition indicated, optimum staging would be as follows:

Desired Percent Removal	Number of Stages
50	1
60	2
75	3
85	3
95	4

Caution, however, must be used when increasing the number of stages to insure that an organic overloading condition is not created (i.e. loading per stage exceeds 6.4 lbs/day/1000 sf).

Step Feeding

At the same influent condition, an analysis of the advantages of step feeding was undertaken. The analysis was undertaken by assigning P and K_s values to each stage to account for the relative quantities of influent flow which are being diverted to downstream stages, but which exhibit kinetics associated with the initial stages of the system. For example, if 75 percent of the flow entered the first stage and 25 percent the second stage

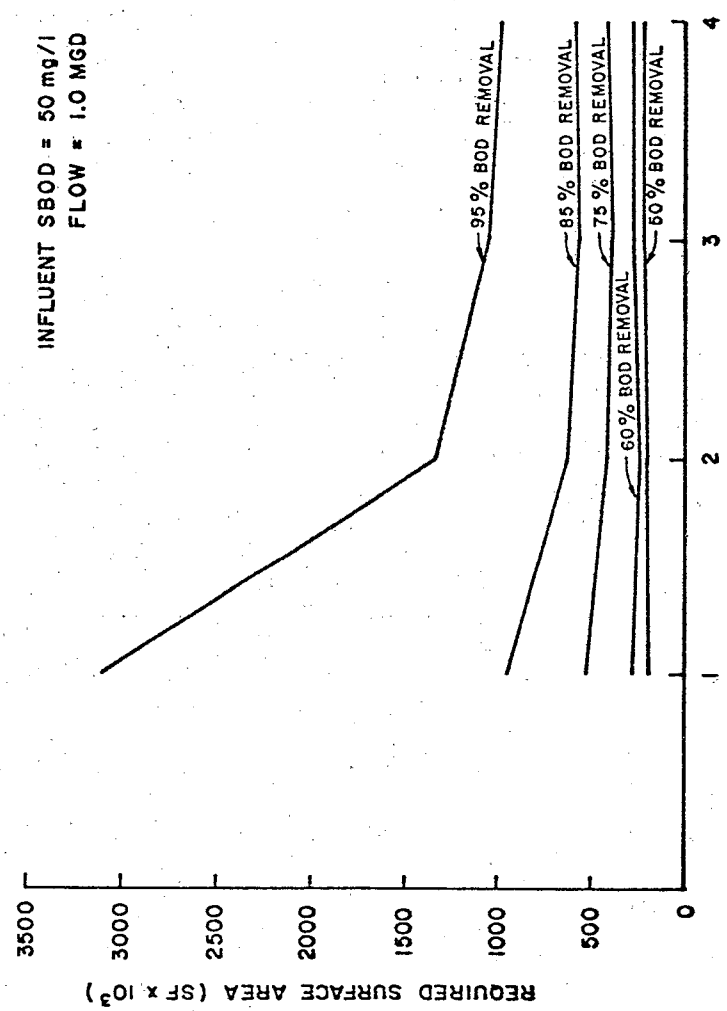


FIGURE 5: EFFECT OF STAGING ON SURFACE AREA REQUIREMENTS

of a 4 stage system, then 75 percent of the flow would experience removal on the basis of running the model consecutively through four stages, with values of P_1K_1, P_2K_2, P_3K_3 and P_4K_4 for the four stages. Twenty-five percent of the flow would experience the removal associated with a three stage system with values of P_1K_1, P_2K_2 and P_3K_3 for the three stages. The final effluent was calculated as a material balance of the two flows. A schematic representation of this analyses is presented in Figure 6. Various combinations of step feeding were examined, with no advantage indicated under any conditions and decreasing efficiencies noted if too great a percentage of flow was diverted to the latter stages.

CONCLUSIONS

1. A limiting organic loading to the first stage of 6.4 lbs/day/1000 sf of BOD (total BOD) was found to be the organic loading beyond which nuisance organism growth and corresponding process performance problems occur.

2. An interstage model was calibrated at organic concentration in a range of 50 - 100 mg/l BOD for systems which were not overloaded. The model illustrates the advantages of staging when high efficiencies are desired. The model did not indicate any advantage to step feeding in systems that are organically underloaded.

3. The design of RBC systems is best undertaken with two independent criteria. The first to establish the desired surface area and staging arrangements, and the second to ensure that no stage is organically overloaded. The organic loading limitation presented here is considered a good design criteria to prevent oxygen limitations. Additional data is required to establish an accurate rational interstage model to predict RBC performance.

NOTATION SYSTEM:

S_{4-3} — PARAMETER (SBOD CONCENTRATION); STEP FEED SEGMENT OF FLOW; STAGE NO.

LEGEND:

Q - % OF TOTAL FLOW
P - MAXIMUM REMOVAL COEFFICIENT, PER STAGE
K - HALF VELOCITY COEFFICIENT, PER STAGE

NOTE:

(1) EACH STAGE = 1/4 TOTAL AREA
(2) INFLUENT SBOD = S_0

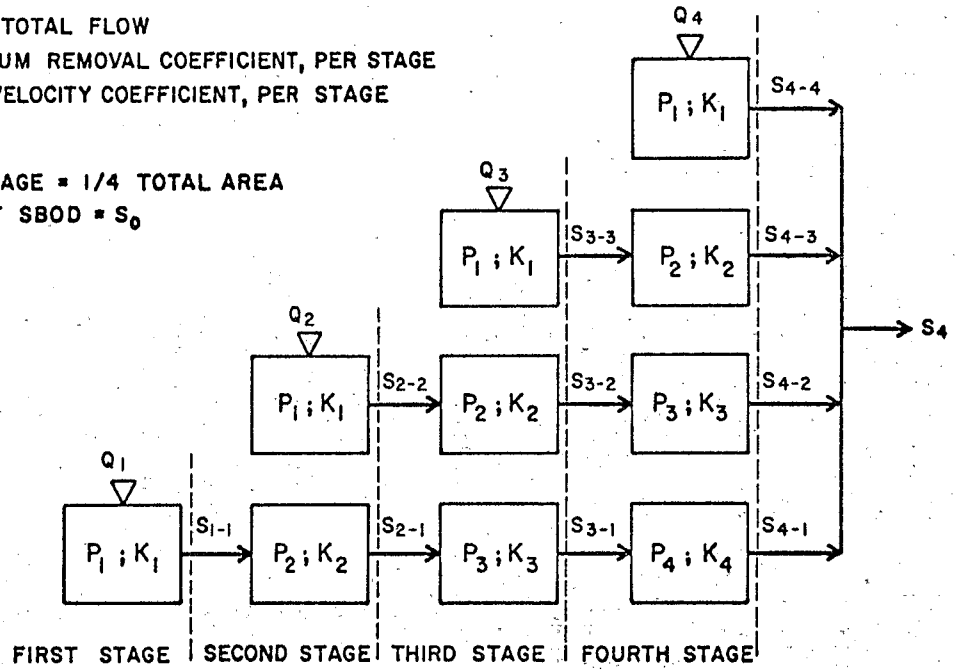


FIGURE 6 : STEP FEED ANALYSIS

REFERENCES:

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COMBINED BIOLOGICAL/CHEMICAL TREATMENT IN RBC-PLANTS.

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INTRODUCTION

Over the last 5 years, there has been an increased use of rotating biological contactors (RBC's) in Norway. Out of a total of about 500 sewage treatment plants in the country, 52 (~ 10%) are RBC-plants.

Since eutrophication is a major pollution problem in the Norwegian waterways, chemical treatment to remove phosphorous is extensively used. Of the 52 RBC-plants, 48 of them are combined biological/chemical plants.

The main objective for establishing chemical treatment in combination with biological treatment in RBC's is of course to remove phosphorous. In addition it is experienced, however, that the addition of a chemical precipitant (normally alum or ferric iron) will improve the effluent quality by coagulation of the fine fraction of biofilm that often is difficult to settle. Therefore the chemical treatment serves a dual purpose.

Chemical treatment may be achieved in combination with RBC's by principally three process designs (fig.1), hereafter named:

- Simultaneous precipitation
- Combined precipitation
- Post precipitation

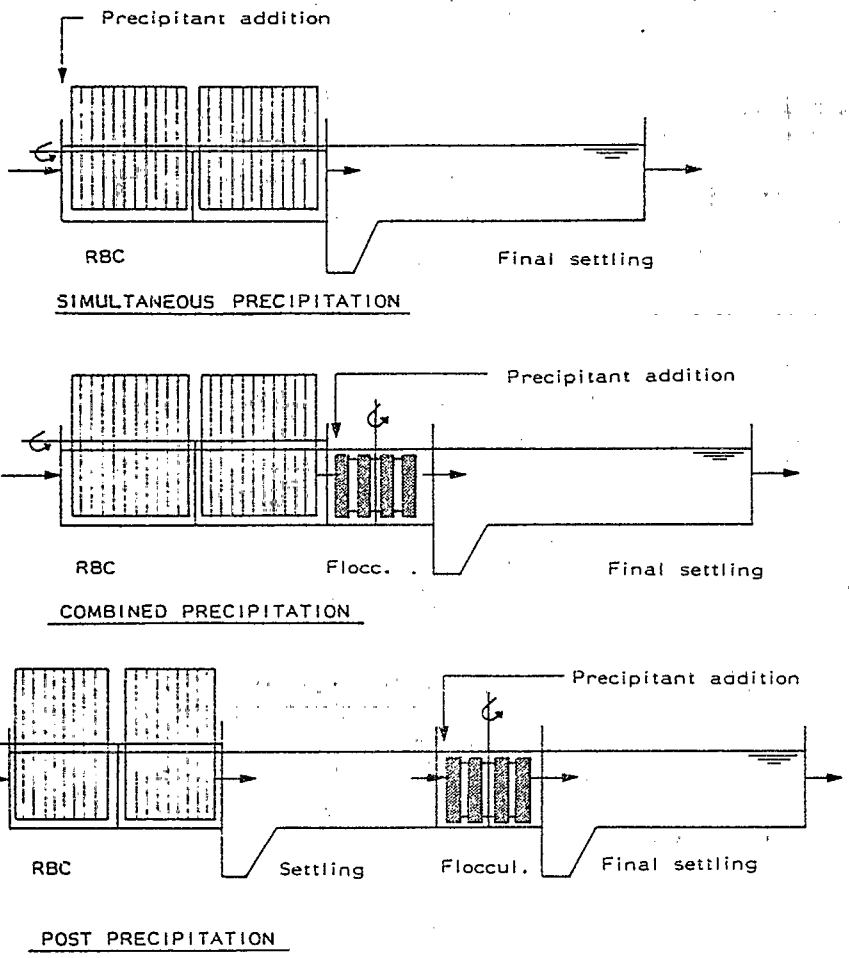

Fig. 1. Different combined biological/chemical treatment processes.

In simultaneous precipitation the precipitant is added to the RBC tank, the precipitation occurs here and the precipitated matter is removed together with the biofilm suspended matter in the following separation unit, normally a sedimentation tank. Since flocculation occurs in the RBC-tank,a flocculation tank between the RBC and the settling unit is normally not included.

Post-precipitation plants consist of biological and chemical treatment completely separated from each other. The RBC has its own settlingtank followed by the chemical step,consisting of chemical mixing,flocculation and separation of flocs.

As will be shown later, the major part of the combined biological/chemical RBC-plants in Norway are designed for combined precipitation. Compared to the traditional post-precipitation plant, the settling unit for the RBC-sludge is here omitted. The precipitant is added to the RBC effluent and the whole suspension (biological and chemical sludge) is then flocculated before combined sludge removal in settling units.

In this paper, an overview over the Norwegian experiences with combined biological/chemical treatment in RBC plants will be given first, and thereafter a special project concerning the comparison of simultaneous precipitation and combined precipitation will be reported.

THE USE OF RBS PLANTS IN NORWAY - AN OVERVIEW.

The general picture of sewage treatment in Norway in Norway is as follows:

- Most of the plants are small. About 50% of the total number of plants (about 500 plants) has a connection of less than 500 personequivalents, and about 93% of the plants are for less than 10.000 person equivalents.
- Chemical precipitation is extensively used either alone of in combination with biological processes.
- All the plants are built-in, either in houses (most of the plants) or in halls blasted into the rocky mountains (the bigger plants). This is so because of the strict labour environment rules and the cold climate during winter.

Based on a questionnaire to the environmental protection authorities in all the counties, information about the RBC plants was gathered.

In table 1 the total number of RBS plants are grouped according to their size and process design.

Table 1. RBS-plants in Norway grouped according to size and process design.

Process design	Size in personequivalents				Total
	> 500	> 500-1000	>1000-2000	> 2000	
Without precipitation	2		1	1	4
Simultaneous precipitation	2				2
Combined precipitation	18	8	11	5	42
Post precipitation	1	1	1	1	4
Total	23	9	13	7	52

It is demonstrated that combined biological/chemical treatment is the normal (48 out of 52 plants) and that combined precipitation is the biological/chemical process design mostly used (42 out of 48 plants).

The principal reason for the popularity of this process is:

- The pollution authorities has accepted this design to give an effluent quality comparable to what is expected from post-precipitation plants.
- While post-precipitation with RBC has not been investment economically competitive with post-precipitation based on activated sludge, combined precipitation has, because of the savings by omitting the RBC settling unit.

There has also been questions whether not simultaneous precipitation also could give results comparable to combined precipitation. If so, the flocculation tanks could also be omitted. Since all the plants are built in, it is obvious that savings in the area of the plants will give considerable savings in the total investment cost. The main objective of the project reported later in this paper was to investigate

this matter.

Effluent quality

The information about effluent quality where not complete for all the 52 plants, partly because many of the plants are so new that the pollution authorities have not started their control program yet and partly because information from the county authorities was incomplete. In table 2 is summarized results of 24 plants where effluent quality has been analyzed on flow proportional samples repeatedly taken over one year. The plants have been divided into two groups ($\gtrless$ 1000 person-equivalents).
All the plants included in table 2 are combined precipitations plants.

Table 2. Mean effluent quality from Norwegian biological/chemical RBC-plants (combined precipitation)

Size group Person equiv.	Tot P ppm	Number of Samples	Number of Plants	BOD_7 ppm	Number of Samples	Number of Plants
< 1000 pe	1,27	42	12	24	35	12
$\geq$ 1000 pe	0,39	70	12	15	70	12

It can be seen that the smaller plants have problems with meeting the effluent standard of phosphorous for this group ($\leq$ 0,8 mg P/ℓ). This is mainly because of operational difficulties with the chemical equipment. In the bigger plants however, which are well operated, the average effluent quality is below the effluent quality standard for bigger post-precipitation plants in Norway($\leq$ 0,5 mg P/ℓ, $\leq$ 20 mg BOD_7/ℓ).

Operational experiences

There are presently 10 different RBC-products represented among the Norwegian RBC-plants. The smaller plants (< 500 pe) are dominated by local products. In the larger plants, three products dominates completely:

- Bio-surf (Aerosurf)
- Envirodisc
- Nova

It is not possible from the data collected to state as to whether any of there products give better effluent quality than the other. In the bigger, well operated plants they all give

effluents quality results that is expected from post-precipitation plants (< 0,5 mg P/ℓ, < 20 mg BOF_7/ℓ).

We have in Norway, as in most countries, I guess, experienced mechanical failures with the RBC plants, and because of this, I can say that the popularity of the RBC's has faded somewhat lately.

Another problem that is experienced, is that nitrification in the RBC reduces the alkalinity so much that it is difficult to maintain sufficiently high pH for chemical precipitation. Alum is normally used as precipitant and pH is then normally 6,0-6,5 in the precipitation step.

RESULT FROM A RBC-PLANT WITH COMBINED PRECIPITATION

For a period of over two years the influent and effluent quality on flow proportional samples have been monitored at Vinstra RBC sewage treatment plant which is operating according to the combined precipitation mode.

The plant is designed for 5100 personequivalents with a design flow of 140 m^3/d. The flow diagram for the plant is shown in fig.2.

The plant receives municipal and dairy wastewater and in addition, external septic sludge is dewatered at the plant. Reject water from this septic handling contributes significantly to the composition of the raw water. The precipitation chemical is aluminium-sulphate of the AVR-quality (a Swedish product consisting of a combination of alum- and ferric sulphate). The dosage of 130-140 mg AVR/ℓ (about 11-12 mg Al/ℓ) is fed flow proportional to the water downstream the biodisc units.

As will be shown later, the average BOD_7 - concentration in the raw water is about 320 g O/m^3 and the average daily flow 1200 m^3/d. Presupposed that a BOD-reduction of 30% will occur in the presettling units, the organic area loading is 19 g $BOD_7/m^2 \cdot d$, which is about the Norwegian design criteria for combined biological/chemical RBC plants. ($\leq$ 20 g $BOD_7/m^2 \cdot d$)

The average treatment results from this plant over the last two years is shown in table 2.

The data in table 2 clearly demonstrate that a well operated combined precipitation RBC-plant can give an effluent quality of at least the same quality as traditional post precipitation plants based on activated sludge can.

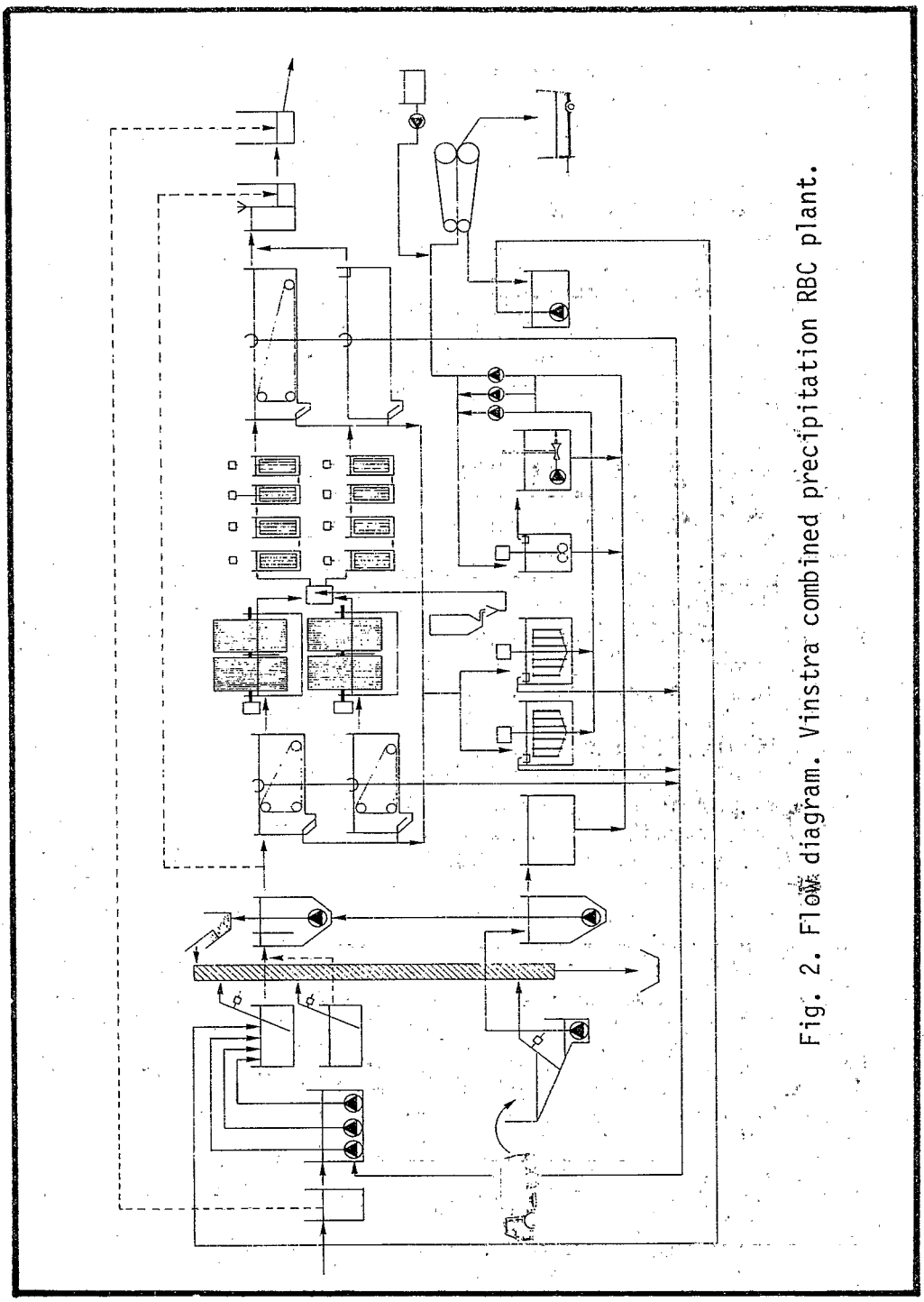

Fig. 2. Flow diagram. Vinstra combined precipitation RBC plant.

Table 2. Average 2-year result from the Vinstra combined precipitation RBC plant.

Parameter	in, g/m^3	out, g/m^3	R %	n
COD	578 ±240	30 ±15	94,8	24
BOD_7	322 ±155	10 ± 4	96,9	23
Total P	8,49 ± 4,2	0,18±0,13	97,9	23
SS	267 ±148	7 ± 4	97,4	24

AN EXPERIMENTAL COMPARISON BETWEEN COMBINED AND SIMULTANEOUS PRECIPITATION.

An experimental investigation was carried out during fall 1981, with the objective of studying combined precipitation at extreme loading and of comparing combined precipitation and simultaneous precipitation. Since the before mentioned Vinstra plant, has two paralell treatment lines, this plant was chosen as experimental site. Two different investigation periods were carried out. In the first period, all of the water was led through only one of the treatment lines after the RBC tanks. In the other investigation period, the flocculation tanks in one of the treatment lines were short - circuited, so that combined and simultaneous precipitation could be investigated in paralell with each other on the same settled raw water.

The two treatment lines are shown in fig.3 where the sampling points are marked. The samples were taken as flow-proportional samples and were analyzed for total and soluble COD, total and soluble PO_4-P, and suspended solids. In addition the secci-depth at the end of the secondary settling tanks were monitored. Since we here are mainly interested in the functioning of the biological/chemical treatment system, the raw water composition described later is that obtained from the outlet of the primary settlers.

The first investigation period lasted for one week and the second for three weeks. The samples were taken daily and analyzed immediately at the plant.

In the second period the point of precipitant addition was changed during the period in order to see if this influenced the results.

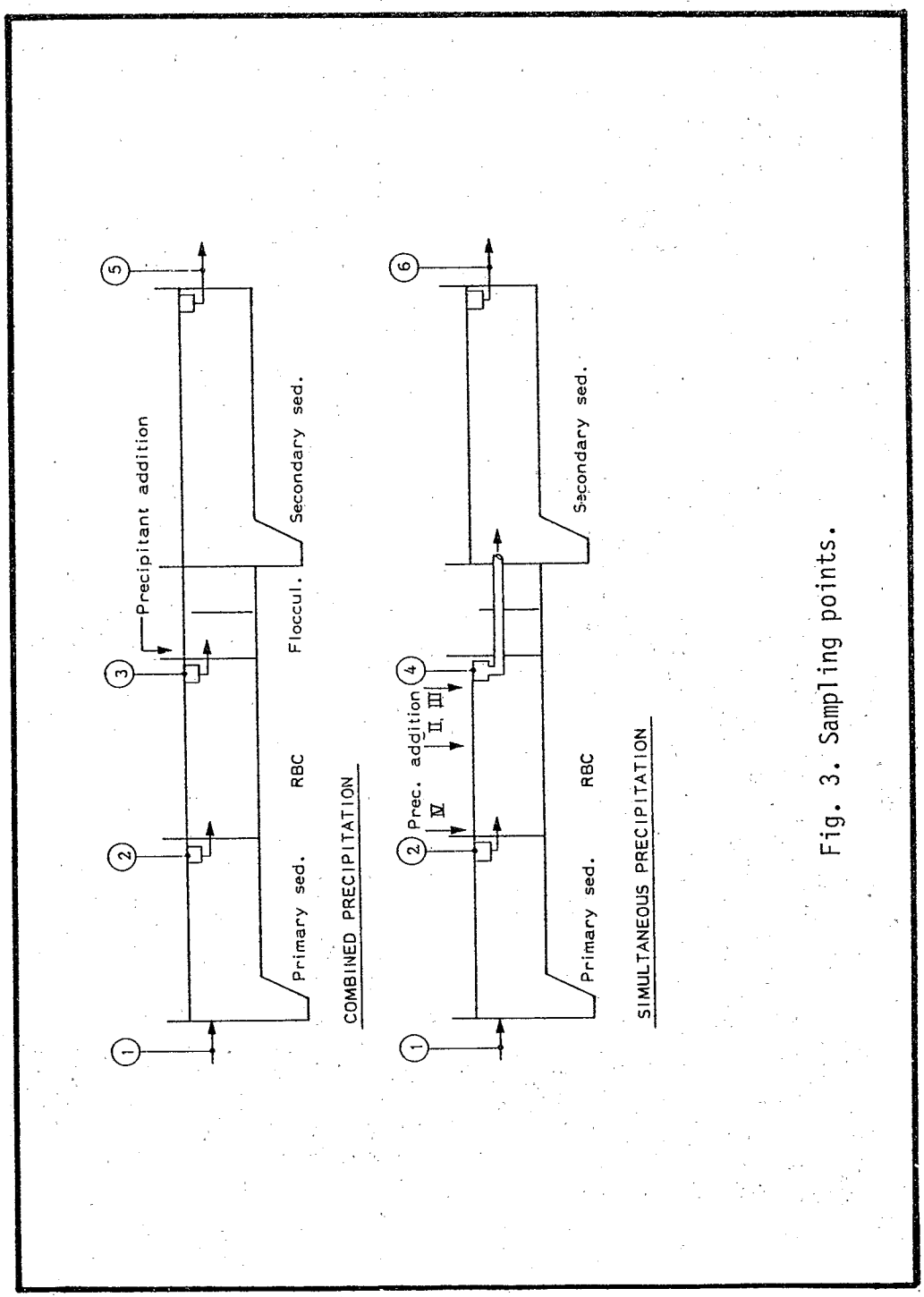

Fig. 3. Sampling points.

In the first period we wanted to investigate how high hydraulic loading influenced the treatment result. As can be seen from table 3 the average daily overflow rate on the final settling unit was 20 $m^3/m^2 \cdot d$, corresponding to an average maximum overflow rate of 1,6 $m^3/m^2 \cdot h$. The Norwegian design criteria for this kind of process is 1,3 $m^3/m^2 \cdot h$.

If so happened that the settled influent in the first period was also very concentrated, partly due to a significant contribution of septic sludge dewatering reject in this period.

Table 3. Organic and hydraulic loadings (1) Calc.values)

	Norw. design criteria	Period 1 Comb.prec.	Period 2 Comb.	Period 2 Simult.
Ave.org.area/load				
$gCOD_{Tot}/m^2 \cdot d$		57	36	36
$gCOD_{sol}/m^2 \cdot d$		26	10	10
$gBOD_{5\ Tot}/m^2 \cdot d$ 1)	18	29	19	19
$gBOD_{5\ sol}/m^2 \cdot d$ 1)		14	5	5
Ave.min.detention time in flocculators min.	20	20	50	0
Ave.max overflow rate in final settlers $m^3/m^2 \cdot h$	1,3	1,6	0,6	0,6

This led to a very high average organic area loading, 57 g $COD/m^2 \cdot d$ based on total COD and 26 g $COD_{sol}/m^2 \cdot d$ based on soluble COD. We have a pretty good knowledge about the correlations between BOD_5 and COD on this water and based on the monitored COD-values, we can calculate the BOD-loadings to have been 29 g $BOD_5/m^2 \cdot d$ based on total BOD_5 and 14 g $BOD_{5\ sol}/m^2 \cdot d$ based on soluble BOD_5.

This means that the plant was both hydraulically and organically overloaded. The organic load was actually considerably over what we had anticipated it to be, namely about the maximum load adviced in the Norwegian design criteria (see table 3).

In the second period the hydraulic loadings were lower partly because the amount of raw water was lower in this period, but mainly because the incoming flow was divided into the two lines. The organic area load was however approximately what was expected, very near the Norwegian design criteria.

In both periods the precipitant dosage was kept at its normal value, 135 g alum/m^3 added flow proportional to the incoming water.

RESULTS AND DISCUSSION

In table 4 are summarized the average treatment result from the two periods.

Table 4. Treatment results.

Parameter		Period 1	Period 2	
		Combined	Combined	Simultaneous
COD_{Tot}	in	561±220	409±104	409±104
	out	31± 10	57± 30	72± 29
$COD_{sol.}$	in	287±137	110± 35	110± 35
	out	21± 9	34± 19	34± 22
PO_4-$P_{tot.}$	in	6,5±2,4	6,7±1,6	6,7±1,6
	out	0,34±0,18	0,39±0,21	0,59±0,13
PO_4-$P_{sol.}$	in	5,6±3,4	2,0±0,56	2,0±0,50
	out	0,02±0,01	0,04±0,03	0,05±0,02
SS	out	13±1	27±9	37±12
Secci depth (cm)		163±33	85±39	50±18

In spite of the extremely high loading in the combined precipitation line in investigation period 1, the plant gave still very good effluent quality similar to what normally was achivied at hydraulically only half the load previously. The COD-values in the effluent corresponds to BOD_5-values of less than 5 g $BOD_{5_{sol}}/m^3$ and less than 10 g $BOD_{5_{tot}}/m^3$.

The phosphate precipitation was also good, precipitation was complete (PO_4-P_{sol} = 0,02 ppm) and even with an average maximum overflow rate as high as 1,6 m/h, the separation of flocs was good, leaving total PO_4-P = 0,34 ppm and suspended solids = 13 ppm in the effluent. The effluent was very clear with secci-depth of 163 cm.

As may be seen from table 4 the effluent guality in period 2 was not so good as in period 1 in either of the treatment lines. Why the concentration of both total and soluble organic matter went up also in the combined precipitation line, we don't know. It is probably however that the extreme organic loading in period 1 had some impact on the biofilm in such a way that the high loading resulted in a thick biofilm that stripped of to a greater extent in period 2 resulting in a decrease in total active biomass.

However, since the main objective in period 2 was to compare the two different processes the absolute treatment result is not so important.

The effluent quality in the combined precipitation line was not bad, however, with average COD-values in the effluent corresponding to BOD_5-values of less than 10 $BOD_{5_{sol}}/m^3$ and less than 20 g $BOD_{5_{tot}}/m^3$.

It was obvious, however, that separation of flocs was worse than in period 1, leaving 27 ppm of suspended solids in the effluent and the secci-depth had fallen to 85 cm. Phosphate removal was still good.

When we compare the results from the two treatment lines, two things are clear:

- The removal of soluble organic matter was equally good in the two lines.
- Separation of flocs was better in the combined precipitation line.

The difference between the two processes lies in the floc separation aspect, as one might expect. This demonstrates the usefulness of the flocculation tanks.

It must be said, however, that the results in the simultaneous precipitation line may have been influenced by:

- the fact that the precipitant addition point was changed over the period.
- the fact that the precipitant dosing equipment for this line was a provisorium which may have given a dosing rate not as reliable as in the other line.

An other investigation period was actually also performed in which all water was led through the simultaneous precipitation line giving a relatively high hydraulic load (ave. max overflow rate 1,2 m/h). The organic loading in this period was relatively low and therefore the results are not included in detail. The results where very similar to the ones obtained in period 2. Towards the end of the hydraulically high loaded simultaneous precipitation period the sludge separation was, however, very good (SS < 10 ppm, PO_4 - P_{tot} < 0,3 ppm and BOD_5 $_{tot}$ < 20 ppm). This proved to us that good treatment may also be obtained by the simultaneous precipitation process.

Based on the results we would, however, advocate that flocculation tanks are used with a detention time of 15-20 min. It is very important that these are constructed so that biofilm settling is avoided.

With regard to precipitation addition point it cannot be stated from this investigation whether this should be done before or after the RBC-tank with combined precipitation. Since only the normal point of addition, after the RBC's was tested here. It may be argued however that precipitant addition before or into the RBC-tank would precipitate some of the soluble organic matter and thus actually reducing the organic load on the RBC. The comparison between simultaneous and combined precipitation did not, however, confirm this, since soluble COD in the effluent was the same in the two lines.

It may also be argued that precipitant addition before the RBC would give less precipitant consumption because the suspended solids concentration are lower here and consequently less precipitant would be consumed in coagulating suspended matter.

In simultaneous precipitation we feel therefore that the most correct precipitant adding point is before the RBC, mainly because this is the point where thorough mixing is easiest to obtain. In the investigation, we did not see much difference in the results, for the different dosing points. But then we kept the dosage constant.

CONCLUSIONS

1. Combined biological/chemical treatment can be obtained in RBC plants by adding precipitant before or into the RBC-tank (simultaneous precipitation), after the RBC-tank and before a flocculation/sedimentation system (combined precipitation) or with separate chemical step downstream the RBC-settling tank (post-precipitation).
2. The experiences from Norwegian combined precipitation plants and from the project reported in this paper, is that excellent treatment results may be obtained by combined precipitation. Since this process has one sludge separation unit less than post-precipitation, combined precipitation is economically favourable compared to post-precipitation.
3. Simultaneous precipitation may also give acceptable effluent quality, but it seems that separation of flocs is better in a combined precipitation system as a result of better flocculation.
4. Adding precipitant (alum) directly to the RBC-tank in simultaneous precipitation does not give any adverse effect on the capability of the biofilm to remove organic matter when pH is kept above pH = 6,0.
5. Good removal of both organic matter and phosphate was obtained with combined precipitation at high organic and hydraulic loadings. A design criteria of average organic loading $\leq$ 20 g $BOD_{7\ tot}/m^2$ d, average max overflow rate in final settler $\leq$ 1,3 $m^3/m^2 \cdot h$ and 20 min detention time in flocculators seem to be acceptable. This is a considerably higher organic loading than can be accepted when chemical treatment is not included.

TREATMENT OF DOMESTIC SEWAGE BY AQUATIC RIBBON SYSTEM

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INTRODUCTION

Fixed-film biological processes have become popular for treating organic wastewaters during the past decade because of their low energy, and possibly low manpower requirements. There are many types of fixed-film biological treatment processes including trickling filters; bio-towers, which are basically trickling filters that use light-weight plastic media instead of gravel as a substrate; rotating biological contactors (RBCs); packed-bed reactors (PBRs); and fluidized-bed reactors (FBRs). The latter three types of fixed-film processes have been successfully implemented for: (1) removal of soluble BOD from wastewaters; (2) nitrification, and (3) denitrification of various wastewaters. These processes generally consist of a fixed-film reactor followed by a liquid-solids separation unit such as a clarifier, a filter, or other special liquid-solids separation unit.

The fixed-film biological processing system reported in this study represents a new concept which combines the fixed-film reactor and the liquid-solids separator into one physical unit. In this new fixed-film system (Aquatic Bio-Ribbon Treatment System), specially designed synthetic ribbons are used as a substrate for microorganisms in a

reactor system to achieve removal of 5-day soluble BOD, nitrification, and denitrification.

SYSTEM DESCRIPTION AND THEORETICAL CONSIDERATIONS

The aquatic ribbon reactor system (Figure 1) consists of a secondary treatment chamber, a nitrification chamber, and a denitrification chamber

in series. In the secondary treatment chamber, zones for distinctly different unit processes are maintained. The upper zone is aerated minimally just to meet all oxygen transfer requirements. The lower zone, which is designed to facilitate sedimentation of sludge particles, is not aerated. Similar to other fixed-film systems, there is no sludge return. The dissolved oxygen (D.O.) level in the secondary treatment chamber is kept above 2.8 mg/l at all times by artificial aeration. Heterotrophic microorganisms are grown on the ribbon surface. The heterotrophs metabolize the organic matter present in the sewage and available oxygen to achieve the conversion of soluble BOD to suspended BOD. Exess biomass growth on the ribbon surface eventually sloughs off and settles into the lower zone of the secondary chamber where it is periodically removed.

The second and the third chambers have an arrangement similar to the first chamber. Both chambers are divided into two zones: the upper zone provides the actual biological treatment process; the lower zone facilitates sludge removal.

The upper zone of the second chamber is aerated to encourage growth of autotrophic nitrifying bacteria (Nitrosomonas and Nitrobactor) on the ribbon surface, but activated nitrifying sludge is not recirculated. Ammonia-nitrogen present in the wastewater becomes the electron donor in the bio-nitrification process and is oxidized to nitrite and eventually nitrate. The D.O. level in the nitrification chamber is maintained at, or above, 1.35 mg/l to promote the necessary nitrification process.

The third chamber is isolated from the outside atmosphere to promote anoxic conditions. This facilitates

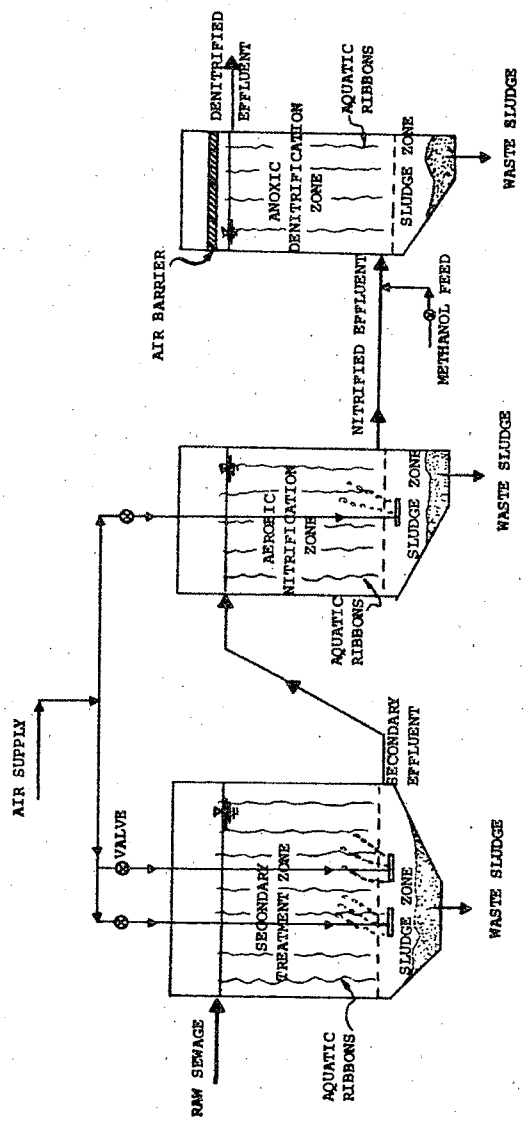

Figure 1. Process Flow Diagram of Aquatic Ribbon Reactor System

growth of denitrifying bacteria on the ribbon surface and thus biological denitrification processes. Denitrifying bacteria which include the genera Pseudomonas, Bacillus, and Achromobactor are responsible for the denitrification process. Nitrate- and nitrite- nitrogen under anoxic conditions become the electron acceptors in the bio-denitrification process.

In the denitrification process, the denitrifiers need soluble carbon to meet the metabolic requirements. As the denitrification process occurs, the carbon source in the wastewater is depleted and becomes a limiting factor. Supplemental carbon is always necessary to enable the denitrification process to continue. In this experiment, carbon was supplemented by adding methanol to the inlet of the denitrification chamber so that the carbon-to-nitrogen ratio was maintained at, or slightly above, 1.10. The empirical methanol feed concentration or requirement can be computed by the following equation:

$$[CH_3OH] = 2.47x[NO_3\text{-}N] + 1.53x[NO_2\text{-}N] + 0.87x[D.O.]$$

where, $[CH_3OH]$ = methanol conc. in wastewater (mg/1)
$[NO_3\text{-}N]$ = nitrate nitrogen conc. (mg NO3-N/1)
$[NO_2\text{-}N]$ = nitrite nitrogen conc. (mg NO_2-N/1)
and $[D.O.]$ = dissolved oxygen conc. (mg/1)

In this study, methanol was manually fed to the third chamber to satisfy the carbon requirement of the denitrification process. A reactor system, which uses part of the influent-soluble BOD as a carbon source for the third chamber, is being studied.

The removal and conversion mechanisms involved in BOD removal, nitrification, and denitrification in an aquatic ribbon system are depicted in Figure 2.

The sewage used in testing the reactor system was primarily of domestic origin. Table I summarizes the composition of the wastewater used in this study. The aquatic ribbon system is currently being studied for its capability to treat tannery wastewater.

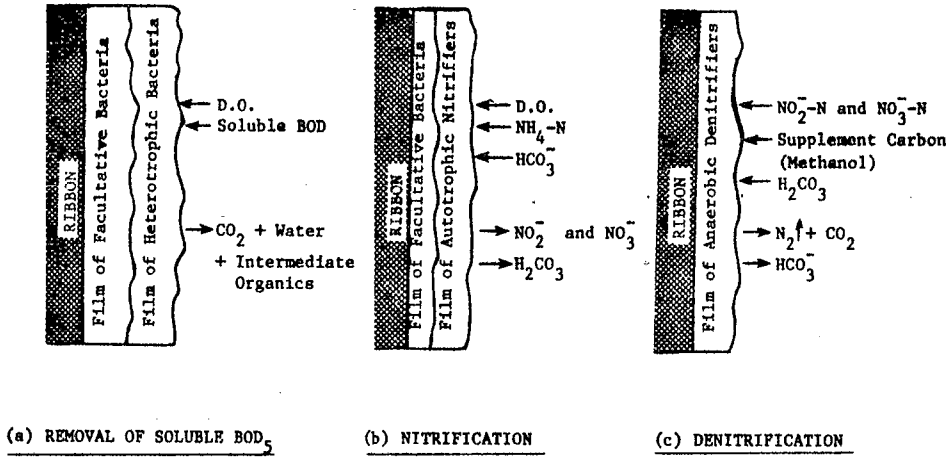

Figure 2. Hypothesized Diagram of Biological Reactions on Aquatic Ribbons

Table I. Characteristics of Wastewater

Parameter	Concentration
pH	7.50 - 7.60
Temperature	26°C ± 2°C
BOD	100 - 248 mg/1
COD	252 - 560 mg/1
BOD/COD Ratio	0.37 - 0.45
TSS	150 - 186 mg/1
VSS	112 - 140 mg/1
Total Nitrogen	29.5 - 34.0 mg/1
Ammonia Nitrogen	20.0 - 24.0 mg/1
Nitrate Nitrogen	0.01 - 0.05 mg/1
Total Phosphorus	8.0 - 12.0 mg/1
Alkalinity (as Calcium Carbonate)	174 - 210 mg/1

The hydraulic detention times in each reactor chamber were varied by changing influent rate. The influent rates tested varied from 7.2 1/day (1.903 gpd) to 21.8 1/day (5.762 gpd). These influent rates result in hydraulic detention times from 8 to 24 hours in the secondary treatment process (First Chamber), from 4 to 12 hours in the nitrification process (Second Chamber), and from 2 to 4 hours in the denitrification process (Third Chamber). Tables II to IV summarize pertinent experimental conditions.

Table II. Experimental Conditions for BOD Removal

Influent Rate (1/day)	Hydraulic Detention Time (Hours)	BOD Loading Rate		D.O. (mg/1)
		Surface (gm/M^2-day)	Volume (gm/M^3-day)	
7.2	24	14.17	248	2.85
10.8	16	13.20	231	2.80
14.4	12	27.43	480	3.20
21.6	8	41.14	720	3.50

Table III. Experimental Conditions for Nitrification

Influent Rate (1/day)	Hydraulic Detention Time (Hours)	Surface Loading of Ammonia-N (gm/M^2-day)	D.O. (mg/1)
7.2	12	1.03	1.35
10.8	8	1.61	1.35
14.4	6	1.80	1.60
21.6	4	3.00	1.80

Table IV. Experimental Conditions for Denitrification

Influent Rate (1/day)	Hydraulic Detention Time (Hours)	Surface Loading of Nitrate-N (mg/M^2-day)	D.O. (mg/1)
7.2	6	1.11	0.15
10.8	4	1.80	0.15
14.4	3	1.94	0.10
21.6	2	2.57	0.40

RESULTS AND DISCUSSIONS

Experimental results for the aquatic ribbon treatment system are summarized in Table V. As shown, soluble 5-day BOD removal was 93.5% at a hydraulic detention time of 24 hours. The soluble 5-day BOD removal efficiency decreased with decreasing hydraulic detention time as evidenced by only an 85.4% BOD removal efficiency when the hydraulic detention time was reduced to 8 hours.

As shown in Table V, nitrification efficiency was 97.6% at a hydraulic detention time of 12 hours. Like BOD removal efficiency, nitrification efficiency also decreased with decreasing hydraulic detention time as evidenced by an 86.5%

Table V. Results of Treatability Study of Domestic Sewage

Influent Rate (1/day)	BOD Removal (%)	BOD Removal Rate	BOD Removal t_d	Nitrification (%)	Nitrification Rate	Nitrification t_d	Denitrification (%)	Denitrification Rate	Denitrification t_d
7.2	93.5	14.17	24	97.6	1.00	12	89.7	1.00	6
10.8	91.2	13.20	16	94.0	1.52	8	88.6	1.59	4
14.4	88.3	27.43	12	86.0	1.70	6	87.1	1.69	3
21.6	85.4	41.14	8	86.5	2.58	4	87.3	2.25	2

Note: Removal rate is expressed in mg/M^2-day
t_d is the hydraulic detention time (hours).

nitrification efficiency when the hydraulic detention time was reduced to 4 hours. Nitrification efficiency was essentially unaffected by the hydraulic detention time, once the hydraulic detention time exceeded 8 hours.

As shown in Table V, denitrification efficiency was 89.7% at a hydraulic detention time of 6 hours. Denitrification efficiency decreased with decreasing hydraulic detention time as shown by an 87.3% denitrification efficiency when the hydraulic detention time was reduced to 2 hours. It should be noted that denitrification efficiency is essentially unaffected by the hydraulic detention time, once the hydraulic detention time exceeded 2 hours.

Figure 3 presents the relationship between soluble BOD removal efficiency and BOD surface loading rate for a given influent BOD concentration. Figure 4 presents the relationship between soluble BOD removal efficiency and BOD volumetric loading rate for a given influent BOD concentration. From these graphs, the following are concluded:

- Soluble BOD removal efficiency increases with decreasing BOD surface loading rate or volumetric loading rate
- Soluble BOD removal efficiency increases with

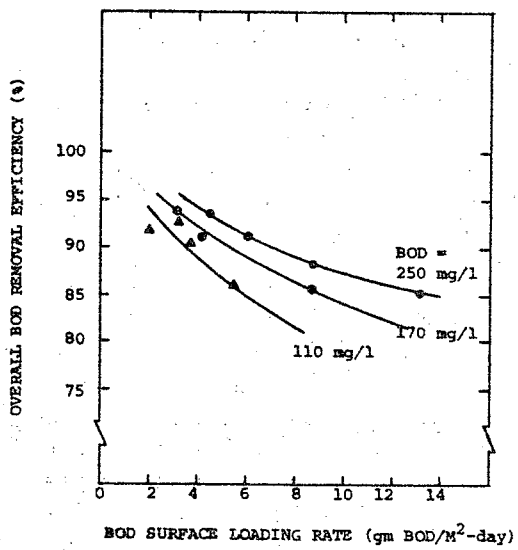

Figure 3. Overall BOD Removal Vs. Surface Loading Rate

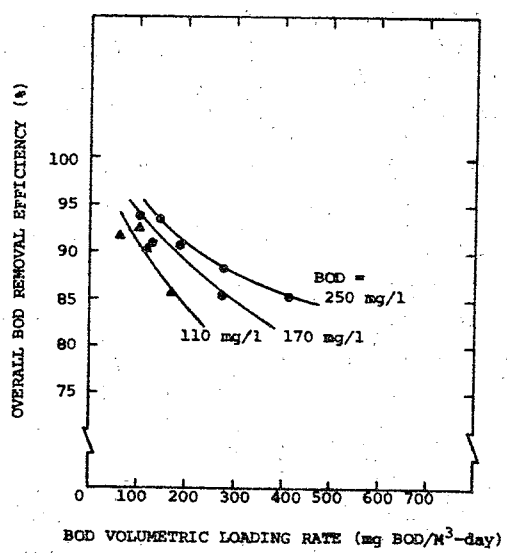

Figure 4. Overall BOD Removal Vs. Volumetric Loading Rate

increasing influent BOD concentration at a given surface, or volumetric loading rate, especially when the loading rates are relatively high.

- The aquatic ribbon treatment system is more efficient at higher influent BOD concentrations, with an upper limit not yet defined.

The overall system BOD removal efficiency, expressed as a function of hydraulic detention time, is presented in Figure 5.

The experimental results related to the nitrification chamber are presented in Figure 6. Nitification process efficiency increases with a decreasing surface loading rate of ammonia-nitrogen. However, the actual nitrification rate (unit mass of ammonia-nitrogen removed per unit surface area and unit time) increases with an increasing ammonia-nitrogen loading rate, and seemingly reaches an assymptotic level. This assymptotic value can not be clearly defined, because it exists beyond the range of the experimental conditions.

The experimental results related to the denitrification chamber are presented in Figure 7. Similar to the nitrification process, the denitrification efficiency is shown to increase with a decreasing surface loading rate of nitrate-nitrogen. The actual denitrification rate (unit mass of nitrate-nitrogen removed per unit surface area and unit time) increases with increasing surface loading rates of nitrate-nitrogen. It is believed that there is an assymptotic value, which again, can not be determined because it lies beyond the range of the experimental conditions.

The nitrification and denitrification efficiencies as a function of hydraulic detention time are presented in Figures 8 and 9, respectively. Figures 8 and 9 also show the actual nitrification and denitrification rates as a function of hydraulic detention time. In brief, the actual nitrification and denitrification rates increase with decreasing hydraulic detention time. The reaction efficiency (whether nitrification or denitrification) increases with decreasing hydraulic detention time. However, the nitrification efficiency drops sharply when the hydraulic detention time is shorter than 4 hours.

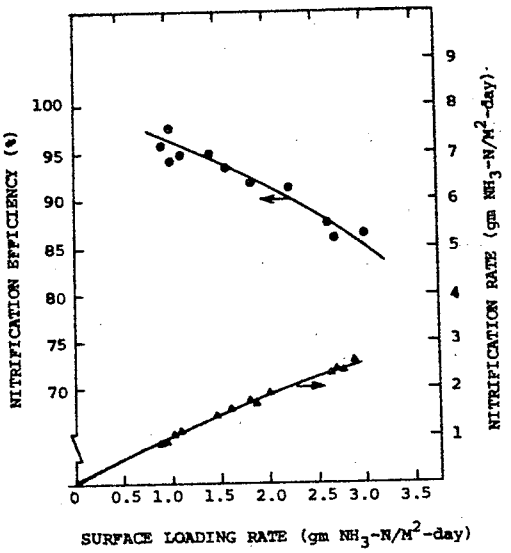

Figure 5. Level of Nitrification Vs. Surface Loading Rate of Ammonia-Nitrogen (NH_3-N)

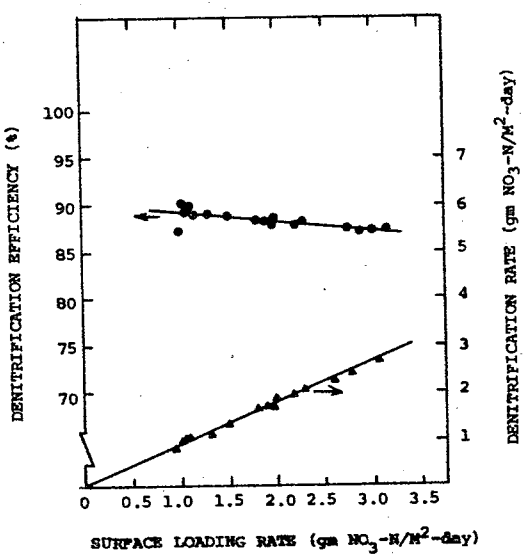

Figure 6. Level of Denitrification Vs. Surface Loading Rate of Nitrate-Nitrogen (NO_3-N)

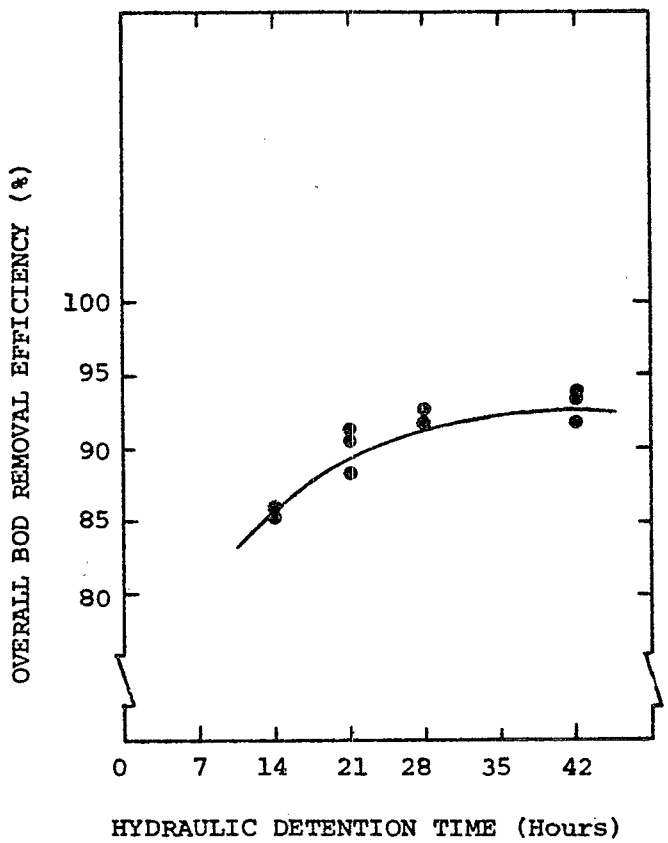

Figure 7. Overall BOD Removal Vs. Detention Time

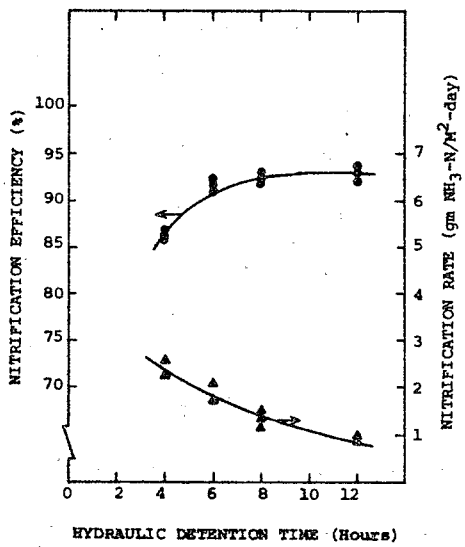

Figure 8. Level of Nitrification Vs Detention Time

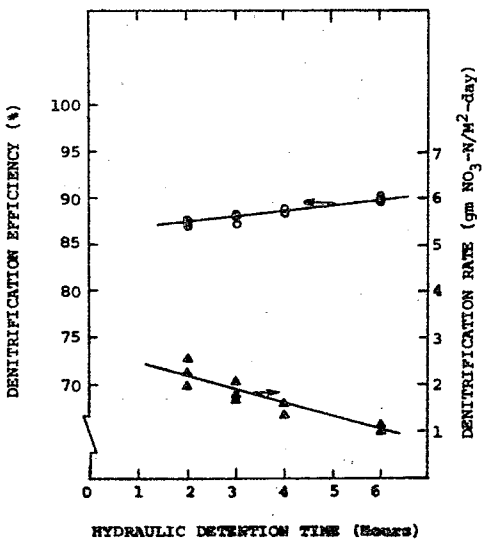

Figure 9. Level of Denitrification Vs. Detention Time

The amount of alkalinity expressed as calcium carbonate consumed per mg/l of ammonia-nitrogen nitrified is presented in Figure 10. The amount of alkalinity generated per mg/l of nitrate-nitrogen denitrified is presented in Figure 11. The net effect on the system chemistry is that every mg/l of nitrogen removed from the wastewater results in an approximately 4.2 mg/l reduction of alkalinity. For domestic sewage, which normally contains 150 to 220 mg/l of alkalinity and 15 to 25 mg/l of ammonia-nitrogen, nitrification and denitrification are not limited by alkalinity availability. If alkalinity availability is anticipated to be a problem, manual addition of alkalinity using lime or calcium carbonate should be considered.

With regard to supplementary carbon for the denitrification process, it was demonstrated that a methanol concentration of approximately 46 mg/l in the wastewater before the denitrification chamber was sufficient for denitrification to take place at 90% efficiency.

It is possible to increase the total surface area of the aquatic ribbons within each chamber of the reactor system to increase the reaction rate, and thus somehow proportionally reduce the hydraulic detention time required. The specific surface area values (i.e., ratio of total surface area and effective liquid volume for each chamber) for the pilot system used in this study are summarized in Table VI along with specific surface areas reported by other researchers (Ref. 1,2,3) using RBC systems for treating domestic or municipal wastewater. It can be seen that the specific surface area used in this study is approximately 5 to 15 times less than those reported by others. Therefore, it appears that the total ribbon surface area could be increased to considerably increase the system capacity and/or reduce the hydraulic detention time required.

COMPARISON WITH OTHER STUDIES

Because an aquatic ribbon fixed-film treatment process is a new concept, no comparable data are available for a comparison study. One can only attemp to make a generic comparison study using the data obtained from RBC systems.

A number of researchers (Ref. 2,4,5,6) have tried to

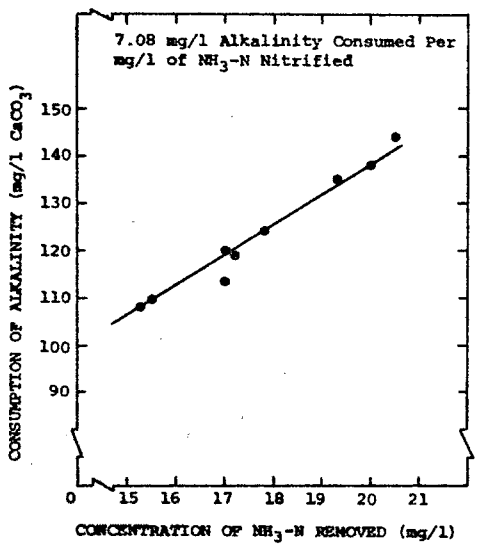

Figure 10. Consumption of Alkalinity as a Function of NH_3-N Nitrified

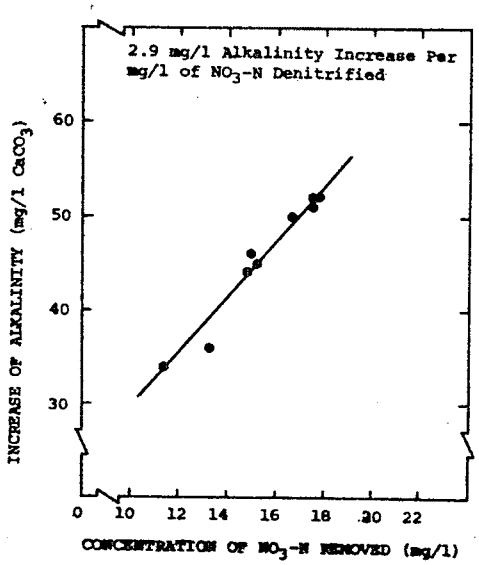

Figure 11. Increase of Alkalinity as a Function of NO_3-N Denitrified

Table VI. Comparison of Specific Surface Area of Aquatic Ribbon System and Rotating Biological Contactor (RBC) Systems

Reference	Secondary Treatment Chamber	Nitrification Chamber	Denitrification Chamber
This Study	18 1/M	40 1/M	70 1/M
Marsh, et al	200 1/M	200 1/M	----
Poon, et al	620 1/M	620 1/M	----
Huang, et al	180 1/M	----	----

establish the relationship between soluble 5-day BOD removal and loading rates. The most commonly used method is to correlate the BOD removal rates with surface loading rates. These relationships by Poon, et al (Curve A, Ref. 2), Lagnese (Curve B, Ref. 4), and Reh (Curve C, Ref. 5), along with the data obtained from this study, are presented in Figure 12. It can be seen in Figure 12 that the soluble 5-day BOD removal rates achieved by the aquatic ribbons system are comparable with RBC systems.

In a nitrification study using a 4-stage RBC system, Marsh et al (Ref. 1) suggested an empirical equation to describe effuent ammonia-nitrogen concentration as a function of influent flow rate, influent ammonia-nitrogen concentration, influent total 5-day BOD concentration, total media surface area available, and wastewater temperature. The equation is:

$$N_e = K \frac{[\ 86.4 x Q x N_o x S_o\ ]}{[\ A x T\ ]}$$

where, N_o = Influent ammonia-nitrogen conc. (mg NH_3-N/1)
N_e = Effluent ammonia-nitrogen conc. (mg NH_3-N/1)
Q = Volumetric flow rate, (cubic meter/sec)
S_o = Influent total 5-day BOD conc. (mg/1)
A = Total media surface area, (square meter)
T = Wastewater temperature, (degree Centigrade)
and K = empirical constant

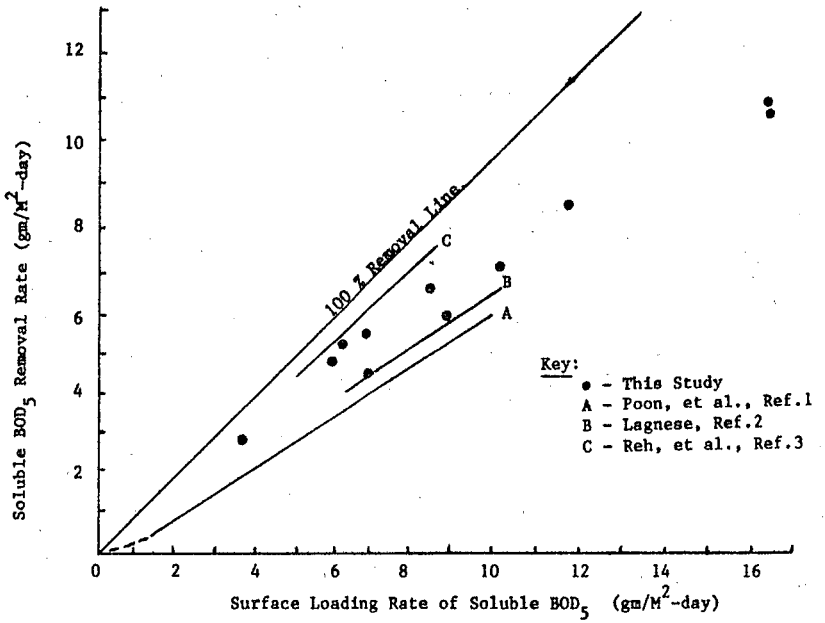

Figure 12. Relationship between Soluble BOD_5 Removal and Surface Loading Rate

According to the above empirical equation, the data obtained from this study and data obtained by Marsh, et al (Ref. 1) were plotted and are presented in Figure 13. Figure 13 indicates that the aquatic ribbon systems provide a level of performance similar to RBC systems with approximately the same slope of K = 15,280.

Poon, et al (Ref. 7), in a nitrification study using a 4-stage RBC system, suggested that the unit ammonia-nitrogen removal rates and surface loading rates are best-fit by a logistic-S curve, expressed by the following equation:

$$R = \frac{R_{max}}{1 + m \cdot e^{b \cdot L}}$$

where, R_{max} = Maximum unit surface nitrification rate
R = Unit nitrification rate
m = Coefficient
b = Coefficient
and L = Surface loading rate of ammonia-nitrogen

Table VII. Comparison of Rmax, b, and m Values for the Nitrification Process with Other Study

Parameter	This Study at 26°C	This Study Adjusted to 11°C	Poon, et al. at 11°C
Temperature	26°C	11°C	11°C
Rmax	3.08	1.37	1.54
m	7.80	7.80	10.28
b	-1.23	-2.76	-2.87

Note: Rmax is expressed in gm/M^2-day.
m is a dimensionless constant.
b is expressed in M^2-day/gm.

Using this logistic-S curve fitting technique, the R_{max}, m, and b values for wastewater temperature at approximately 26 degree centigrade and temperature-adjusted

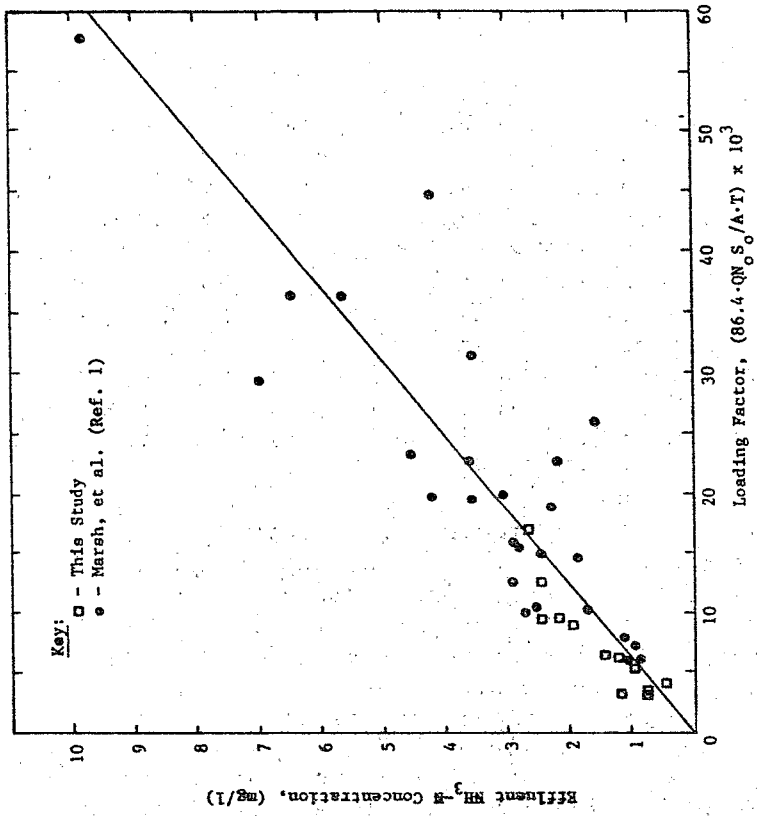

Figure 13. Effluent NH_3-N Concentration Versus Loading Factor

R_{max}, m, and b values from the data obtained from this study are presented in Table VII. Values reported by Poon, et al are also listed in Table VII. From Table VII, it can be seen that the aquatic ribbon fixed-film treatment process is similar to RBC systems in level of treatment performance, and can be predicted fairly accurately by the logistic-S curve as suggested for RBC systems.

CONCLUSIONS

The aquatic-ribbon fixed-film biological treatment process discussed in this paper is a newly developed treatment technology, which is capable of achieving removal of soluble BOD, nitrification, and denitrification from domestic or municpal wastewaters at a level similar to conventional rotating biological contactors (RBCs). Under the experimental conditions in this study, the pilot aquatic ribbon treatment system was capable of removing more than 91% of the total 5-day BOD at a hydraulic detention of 16 hours; providing 94% nitrification at a hydraulic detention time of 8 hours; and achieving 87% denitrification at a hydraulic detention time of 2 hours.

The combination of the liquid-solids separation process with the aquatic ribbon reactor into one physical unit is a unique feature of the aquatic ribbon fixed-film biological treatment system. This feature effectively eliminates the need for a separate clarifier and thus should result in a cost-savings for aquatic-ribbon systems compared to conventional RBC systems. Because the liquid-solids separation zones in the secondary treatment and nitrification chambers are connected to the upper reaction zones, aerobic conditions can be maintained at all times. This reduces the potential of rising sludge and bulking problems which are commonly encountered, when septic conditions occur in the bottom sludge of a conventional clarifier.

There are areas for further improvements to the aquatic ribbon treatment system. The pilot system used in this study has an effecive specific surface area of approximately 5 to 15 times less than typical values for conventional RBC systems. Consequently, there appears to be a great potential for aquatic ribbon systems to achieve a level

of performance equal to or better than conventional RBC systems cost-effectively.

Aquatic ribbon biological treatment systems are a low-technology, low-energy alternative to other conventional treatment technologies. Aquatic-ribbon systems can be easily incorporated into existing lagoons and activated sludge treatment plants without significant process modifications to improve levels of treatment and reduce energy consumption.

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ACTIVATED FIXED FILM BIOSYSTEMS IN WASTEWATER TREATMENT

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I. OVERVIEW

Historically, waste containing organic materials have been subjected to biological treatment processes to reduce the impact of the waste on receiving environment. The most widely used concept in urban areas has been the activated sludge process, a fluid bed system. With waste of increasing complexity from municipalities due to industrial influxes and changing life styles, the basic biological waste treatment process has been severely stressed in many locations to perform satisfactorily. The activated sludge system was encouraged in the 1970s due to its efficiency of BOD removal as compared to the fixed film biological reactors, i.e., trickling filters. The future of new construction of activated sludge systems or wastewater treatment in general is clouded due to the de-emphasis by the present administration in Washington. While the need for environmental improvement and management remains, federal funds for the construction have been severely eliminated. New systems will have to be justified on the basis of savings or on the basis of least cost to the municipality. With the emphasis on local financing of wastewater systems, and the need for stable operation due to shock loading and variable levels of

of toxic materials, the fixed film biological system again appears to have certain advantages over the activated sludge process.

The traditional fixed film reactor has been the "trickling filter" with either a stone or plastic media. These units were originally installed due to their simplicity of operation and their low energy requirements. Their limited removal efficiency and susceptibility to shock loading were accepted as trade-offs for their advantages. The introduction of the synthetic or plastic medium greatly improved the efficiency of operation of the trickling filter but at the sacrifice of more sophisticated operation and more energy input. Directionally, advances were made in fixed film reactors with the development of rotating biological contactors to provide a low energy input system but also a system which has improved operational characteristics. As indicated by Benjes (1), the manufacturers of rotating biological disks claim that the system offers the following advantages as compared to conventional fluid bed systems.

1. Simpler operations
2. Capability of meeting a 10/10 effluent standard without subsequent treatment
3. Final clarifier under flow concentration from 2 to 3 percent
4. Lower energy requirements per pound BOD removed

The standard rotating disk operation is a once through flow process with no recycling involved. Obviously, this presents a significant operational advantage over other processes. The capability of the biological system to meet oxygen demand requirements in the first stages of the process has been questioned. Zero dissolved oxygen will probably occur and potential odors will result. The claim of achieving medium and low effluent standards without additional treatment is probably true only of lightly loaded systems. Excellent quality has been shown to be achievable of operating plants; however, the ability of the system to reach extremely low levels of effluent BOD and suspended solids is questionable. The final underflow concentration does reach the 2 to 3 percent level which offers an advantage in the sludge handling facilities. The primary advantage claimed for the system is the lower power requirement. Although this is probably true, based upon pilot plant and operating studies, for reasonable effluent criteria (30/30), this does not appear to be a valid claim

as the effluent concentrations become lower. The power requirements appear to approach those for activated sludge systems as these lower effluent concentrations are required.

A second type of fixed film system has also been advocated by certain manufacturers and engineers as a competitor for the conventional activated sludge system. As described by Richter (2), the activated bio filter (ABF) system is actually a combination of a fixed film reactor and a fluid bed system. Biological solids which have been clarified from the fluid bed system are recycled back to the top of a fixed film reactor and allowed to flow through the reactor into the fluid bed unit. This combination of reactors has been utilized in potato processing waste in Idaho with very satisfactory results. The combination of the two processes is reported to provide a very stable system when receiving highly variable influent loads and to provide a very rapidly settling biological floc. Benjes (1) also evaluated the ABF system as a competitor to the traditional activated sludge unit. Although this analysis (1) was not based upon any one location or operation, the same general advantages were discussed, i.e., for reliable operation under varying loads, lower capital and operating costs, and simpler operation. It was suggested that each of the claims had to be evaluated on a site specific basis and should not be accepted as generalizations.

Research performed at Memphis State University over the past several years has advanced the information available for both rotating biological contactors and the ABF process with not only city of Memphis municipal wastewater but also synthetic wastewater utilizing glucose substrate. The results of the MSU investigations will be presented in the following paragraphs for both pilot plant field studies and laboratory studies. Under a research grant from the city of Memphis, Division Public Works, pilot plant studies were conducted over a one-year period at the T.E. Maxson wastewater facility in Memphis. Comparable results were obtained on an ABF system, a trickling filter utilizing plastic media, and a conventional contact stabilization activated sludge systems. A cost effective analysis of the proposed expansion of the T.E. Maxson facility prepared by Black and Vetch engineering consultants in Kansas City, Missouri, indicated that not only would the ABF system present a lower capital cost alternative but would also be less energy intensive and actually reduce the energy consumption of the existing plant. The laboratory studies utilizing rotating

contactors constructed of wood followed by a fluid bed aeration system have indicated that the process can be operated at high F/M ratios with reasonably consistent performance. The activated RBC system has also been shown to be resistant to toxic loads through the development of a balanced eco system to degrade resistant pesticides.

II. ABF PILOT STUDIES

The ABF pilot plant, as shown in Figure 1, received wastewater from the aerated grit chamber of the T.E. Maxson wastewater treatment plant. The degrited wastewater passed through a circular primary clarification basin then into a mixing sump where the clarified effluent was mixed with return sludge from the secondary clarifier and bottoms from the biotower and pumped to the top of the biotower itself. From the biotower, a portion of the flow which was not recirculated was transferred to a short term complete mixed aeration basin and then through a secondary clarifer where the biomass was separated and either recycled or wasted to an aerobic digestor. A similar unit was operated at the city of Memphis North Wastewater Treatment Plant. The biofilter was obtained from the Neptune Microfloc Company on a loan basis for the pilot plant evaluation. The filter, a 24-foot high, 4-foot square unit, contained 21 feet of horizontal wood medium consisting of wooden slats on one-inch horizontal spacing with flights arranged every six inches. The operation in the biotower was controlled by the organic loading (pounds of BOD per cubic foot of medium) as well as hydraulic loading (gallons per square foot of medium). The effluent from the biotower contained high concentrations of biological solids, both those entering the tower and those which had fallen from the horizontal wooden medium. The aeration basin was also provided by Neptune Microfloc and was considered as an integral part of the fixed film treatment concept. The flow from the biotower was subjected to an aeration period of approximately two to four hours in order to stablize the biological solids and allow for any additional removal of soluble BOD. Air was supplied to the system from the main air supply of the treatment plant through diffusers (coarse bubble) located in the bottom of the aeration basin. The effluent from the aeration basin was clarified in a seven-foot diameter secondary clarifer. This unit was not adequate for the flows placed through the system as indicated by the surface over flow rate exceeding

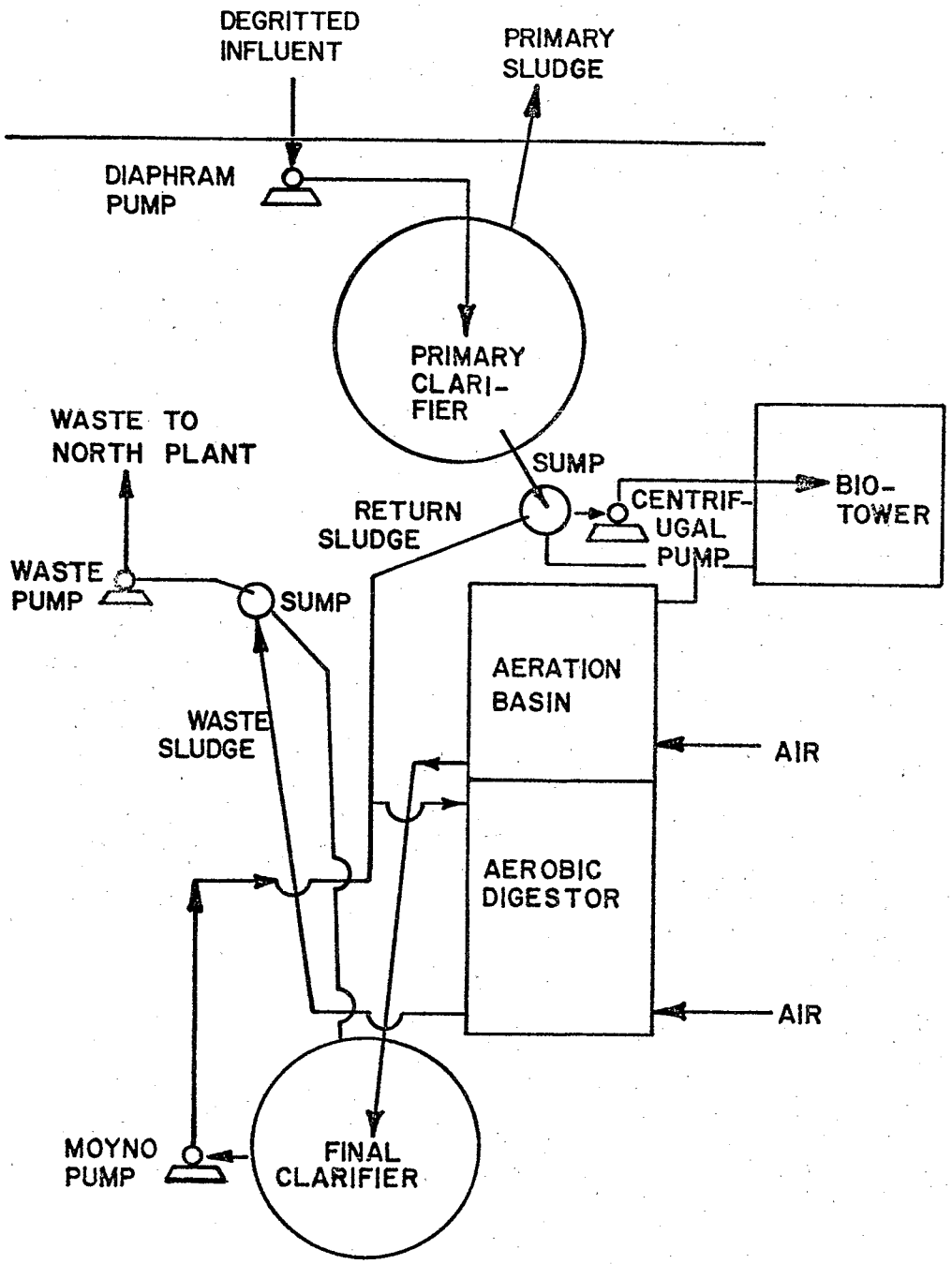


Figure 1

ABF PILOT PLANT LAYOUT

1700 gallons per day per square foot at times. The higher than reasonable overflow rates caused the lower quality effluent than would have been realized through an adequately designed and sized clarifier.

The operational theory of the ABF system is relatively simple but yet not normally experienced in waste treatment systems. By recycling the underflow from the secondary clarifier to the top of the biotower, a high microbial solids level is achieved within the tower itself. The ABF system utilizes a wooden horizontal medium as opposed to a plastic medium. The horizontal medium not only allows the microbial solids growing on the media to remain active longer because of the moisture content of the wood medium, but also provides finer droplet formation within the tower due to the flow pattern around the horizontal wooden boards. The flow rate through the system was initially set at 1.5 gallons per minute per square foot wetting rate on the tower. This proved to be an unstable operation condition due to uneven sloughing of solids from the tower. The unit was operated slightly over three weeks in this mode and then the wetting rate increased to 2 gallons per minute per square foot at which point a uniform, constant sloughing rate was achieved. At this wetting rate the raw wastewater flow into the tower was 16 gallons per minute with a recycle of return activated sludge of 7 gallons per minute and a recycle from the tower of 9 gallons per minute providing the 2 gallon per minute per square foot wetting rate. Hourly grab samples and 24 hour composite samples were utilized to evaluate the performance of the system. The evaluation parameter in the biotower was the loading in terms of pounds of BOD per 1000 cubic feet. It is believed that the fixed film reactor (ABF tower) operated as both an absorption medium for collodial solids and colloidal BOD and as a biological oxidation region due to the high surface area to which the wastewater is exposed. The aeration basin is utilized to allow time for microbial stabilization of the remaining BOD coming from the bottom of the biotower. The dissolved oxygen level and detention time in the aeration basin were varied in this study to evaluate the minimum and maximum values which could be utilized. Also, the loading rate across the system in terms of a system food to micro-organism ratio was observed and correlated with percent removal as will be discussed in later paragraphs. The concept of a system F/M ratio is a valid one for this type of a biological system.

Although the influent was highly variable in BOD and suspended solids, an equalization basin was not provided ahead of the ABF tower. The reason for this exclusion was to provide a severe evaluation or test of the ABF unit by itself to equalize load fluctuations or conversely to absorb shock loads. The assumption was that if the system performed adequately without an equalization basin, it would present many advantages to the city of Memphis or others who were investigating this type of unit as a retro-fit to an existing plant.

Beginning in March and April, a B.F. Goodrich plastic media filter was replaced with the ABF tower supplied by the Neptune Microfloc Corporation of Corvallis, Oregon, containing horizontal wood media. The operational mode was changed because of the nature of the ABF system. Because the biotower was operated at a relatively constant hydraulic loading, wide fluctuations in the organic loading in terms of pounds per thousand cubic foot per day of BOD were experienced. The biotower, because it contained a high population of biological solids on horizontal medium, was also monitored for parameters related to a normal aeration system, i.e., oxygen uptake rates and sludge volume indices. The biotower performed extremely well and much better than was originally anticipated when it was installed. As shown in Figure 2, the organic loading varied from a high of 452 pounds per day per thousand cubic feet to a low of less than 110 pounds per day per thousand cubic feet. Even with this wide fluctuation in loading, the biotower removed a consistent level of soluble BOD. The biotower functions very similarly to the high rate plastic media filter in the sense that a significant soluble BOD removal is anticipated. The effluent from the biotower contains high levels of active biological solids which render a total BOD analysis not applicable. The oxygen uptake rates at the bottom of the biotower were relatively high when compared to the pilot plant complete mix aeration basin following the biotower (see Figure 3). The solids settled reasonably well as indicated by the sludge volume index values. Several studies were performed by taking hourly samples of the influent and effluent from the biotower to evaluate the ability of the biotower to absorb shock organic loadings. These studies are summarized in Table 1. As can be seen by analysis of the data in this table, the soluble removal across the tower was generally greater than 80% and often times reached as high as 97%.

FIGURE 2.

BIO-TOWER PERFORMANCE

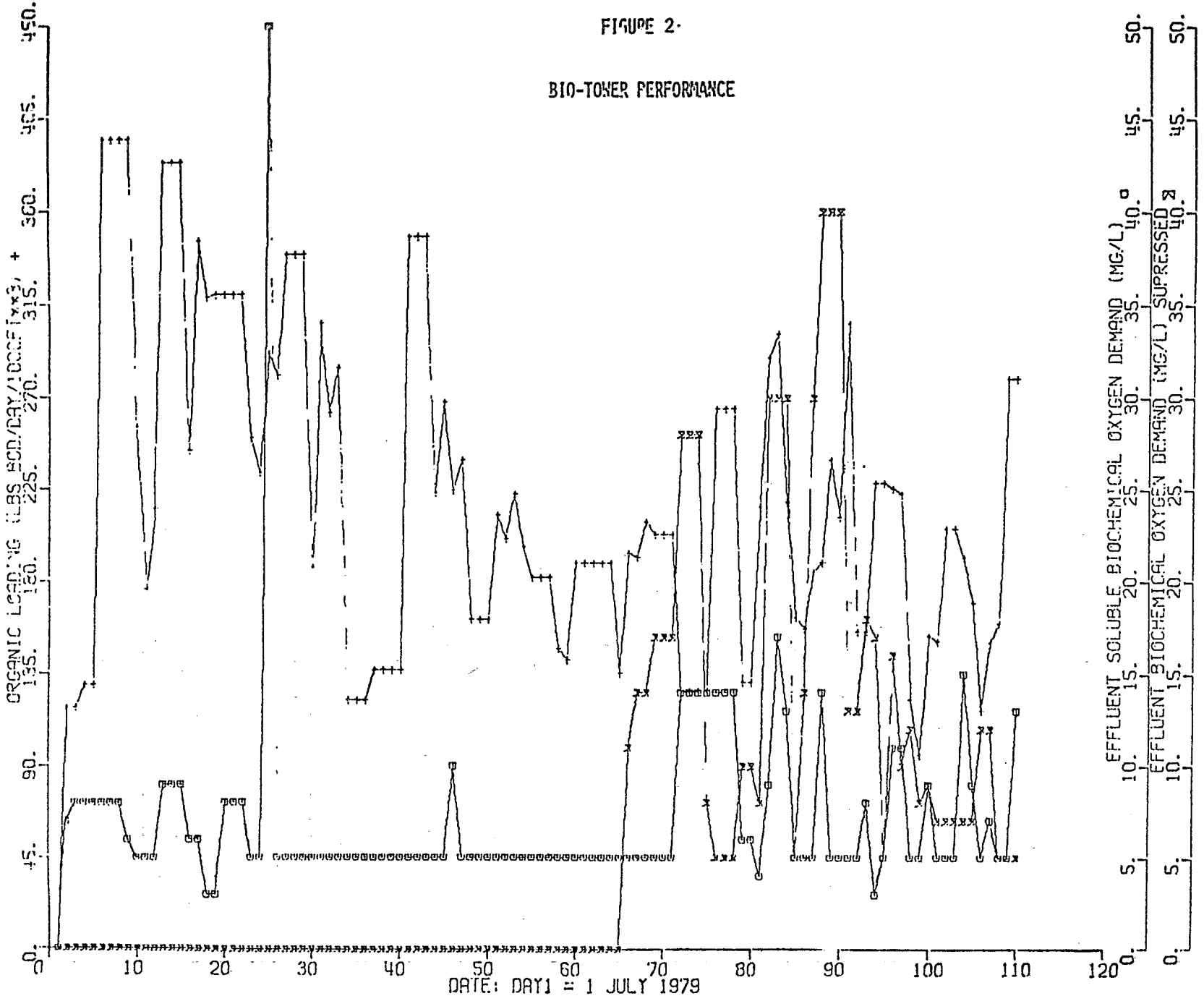


FIGURE 3

OXYGEN UPTAKE RATES IN BIO-TOWER
AND AERATION BASIN

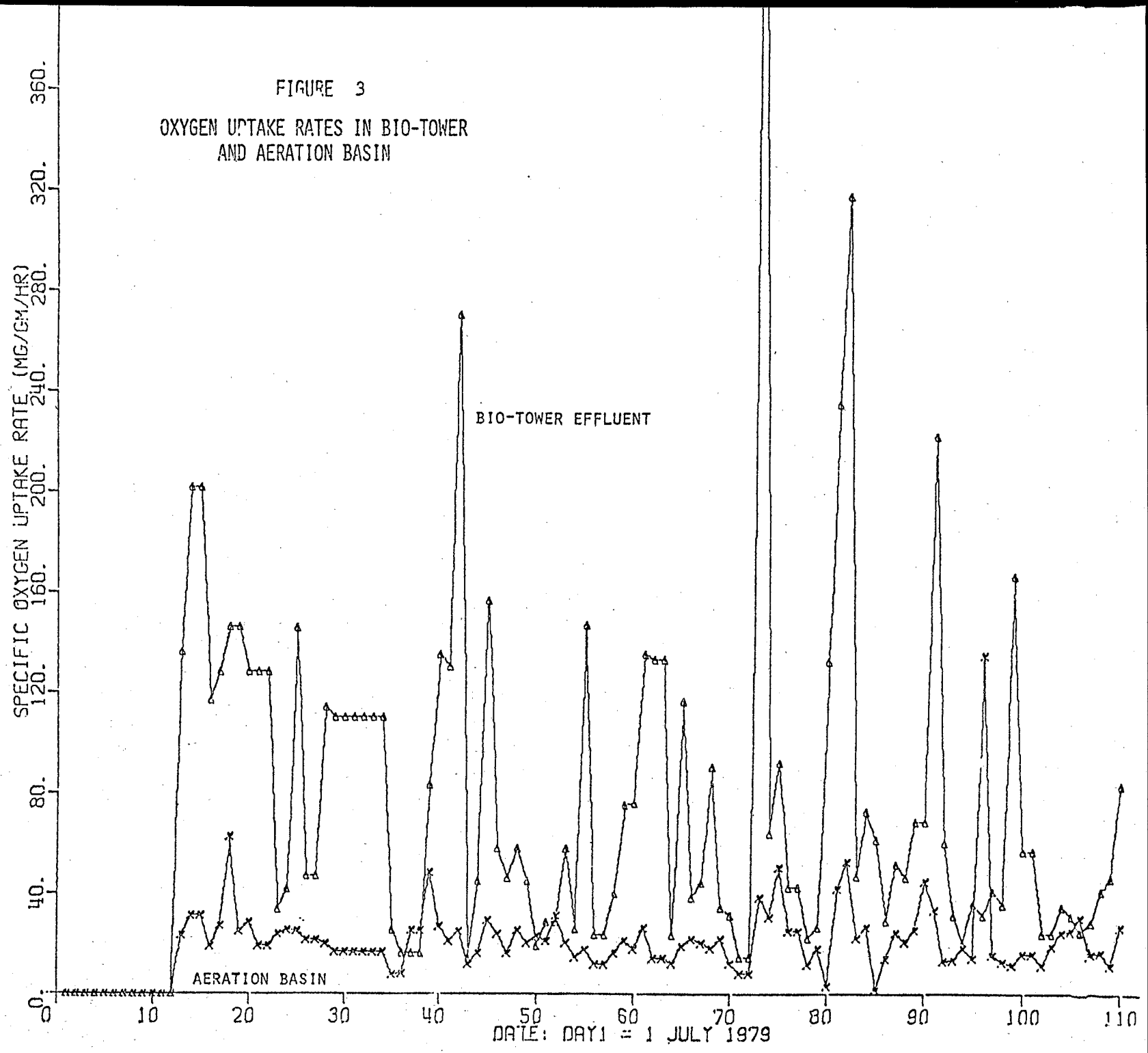


TABLE 1

Bio-tower Evaluation

Soluble BOD Removal

Soluble BOD

Sample #	Influent	Bio-Tower Eff.	% Rem.
1	270	68	75
2	294	42	86
3	274	59	79
4	276	90	69
5	270	45	83
6	279	42	85
7	363	50	86
8	330	51	85
9	310	40	87
10	315	9	97
11	282	24	92

Note: Flow conditions were 14 gpm primary effluent, 10 gpm bio-tower recycle, and 7 gpm return sludge.

TABLE 2

Effect of Aeration Time
on Soluble BOD Removal

Elapsed Time (hours)	10-17-79 Soluble BOD Remaining	10-2-79 Soluble BOD Remaining
0	98	11
1	24	1
2	16	6
3	8	1
4	9	1
5	5	3
6	13	5
7	–	5
8	–	5

The amount of aeration which was required to stabilize the underflow from the biotower was an unknown entity. A laboratory study was performed with the flow from the tower bottom to determine the optimum aeration time. This study was performed by pulling samples of the biotower bottom flow, aerating it for a prolonged period of time while pulling samples of the mixed liquor at various time increments. Analyses of laboratory and field studies indicated that an aeration time of less than four hours and probably less than three hours would be adequate to remove most of the carbonaceous BOD (see Table 2). As shown by an analysis of the data in Figure 4, the aeration basin was operated at varying mixed liquor suspended solids levels with the mean cell residence time at or about 4½ to 5 days. The mean cell residence time was calculated based on the amount of mixed liquor solids in the aeration basin. An alternate procedure using total solids inventory in the system was not utilized. The F/M level was an arbitrary point as far as the study was concerned but it tended to provide an indication of the stability of the system. The oxygen uptake rate in the aeration basin was relatively low (around 40 mg/l per hour) and stable even with a highly variable uptake ratio in the tower bottom (see Figure 3). The low oxygen uptake rate was an indication of the low level of soluble BOD entering the system. With the ABF process, it is almost inappropriate to speak of the biotower without speaking of the activated sludge portion. Using a systems analysis approach where the system considers the total load of the biotower as the food and the microorganisms in the aeration basin as the amount of microbes, a range of system F/M between .3 to greater than 2 was observed. The total system performance was found to be less influenced by the loading on the biotower in terms of pounds per day per thousand cubic feet than on the detention time in the aeration basin.

Based on an assumed 80% removal of soluble BOD across the biotower, the soluble loading onto the aeration basin in terms of pounds per thousand foot of aeration volume per day became relatively low. The aeration time proved to be a critical factor in the level of BOD in the effluent. It is normally recognized that the operation of a system is satisfactory when the soluble BOD level is consistently less than 10 mg/l in the effluent. The total BOD in the effluent of the aeration basin was at or above 30 mg/l for most of the test period. It was determined toward the end of the test program that nitrification was occurring in the effluent

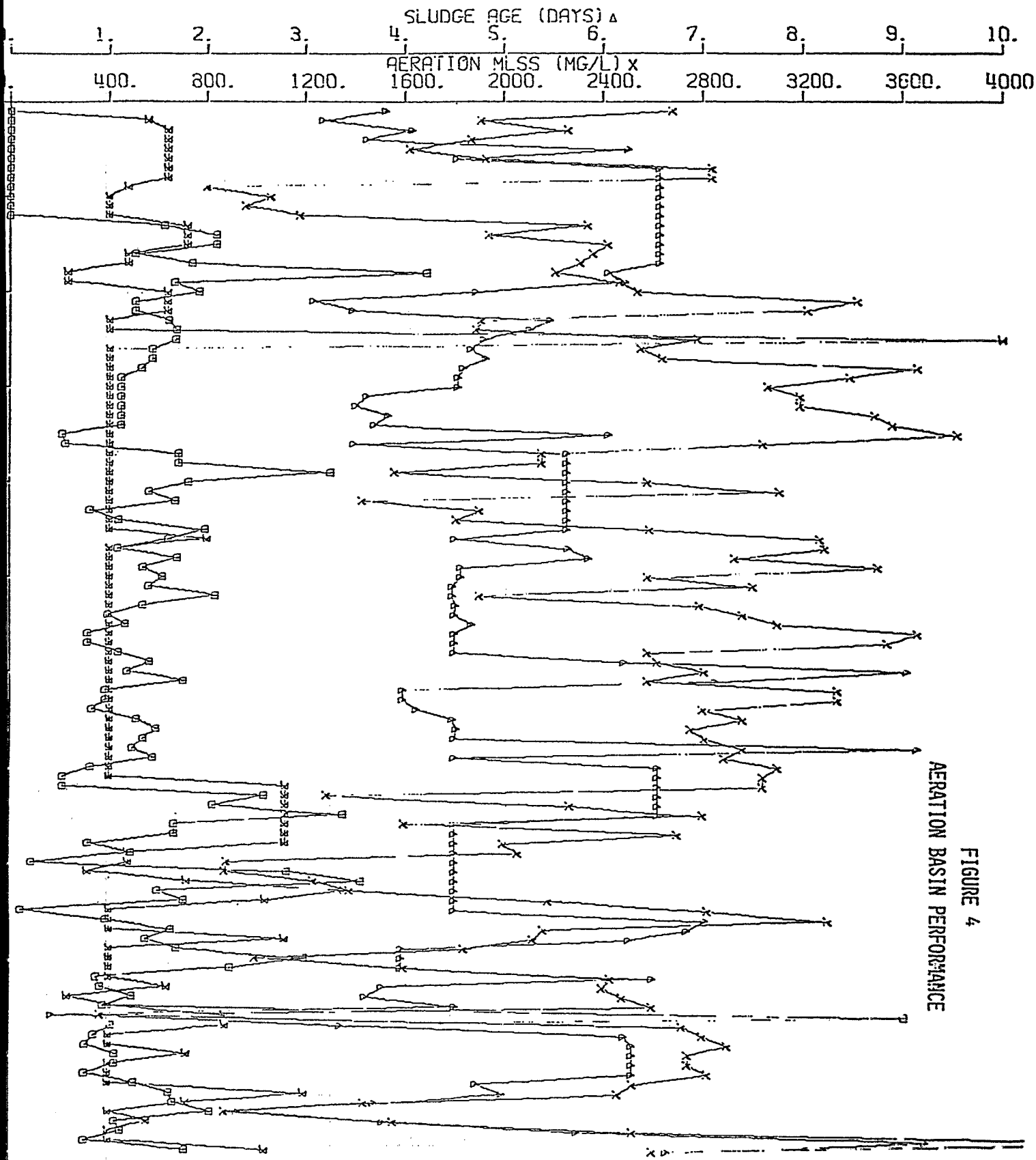


FIGURE 4
AERATION BASIN PERFORMANCE

samples and steps were taken to alter laboratory procedures to compensate for the nitrogenous BOD. When this compensation was made, the pilot plant unit achieved the desired effluent quality of less than 30 mg/l BOD and suspended solids.

III. LABORATORY ARBC STUDIES

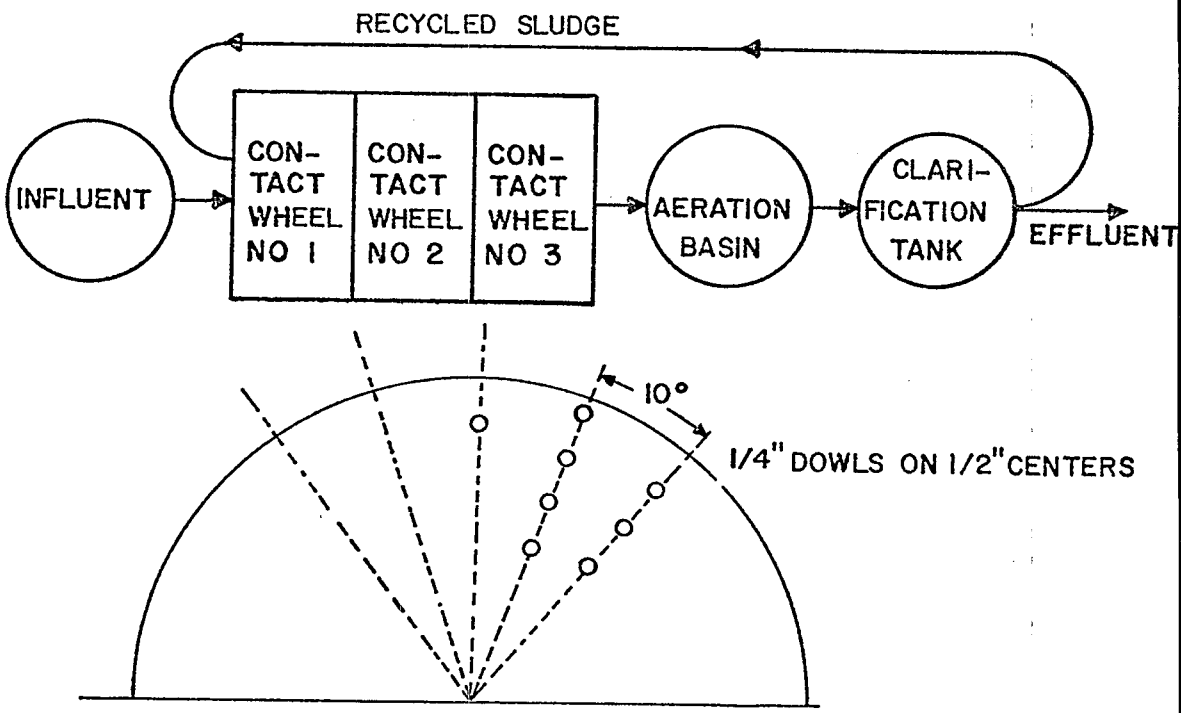
To advance further the observations made with the ABF system, a laboratory fixed film unit was developed as shown in Figure 5. Using wooden dowels for the support media, the rotating contactors were exposed to a mixture of wastewater and underflow solids from the secondary clarifier. Normal treatment parameters (BOD, COD, SS) were measured on the influent and effluent of the system. Mixed liquor suspended solids were also determined for each wheel and the aeration basin. Influent to the North Treatment Plant as well as a synthetic wastewater using glucose as a carbon source were used in the laboratory studies. Shock load conditions (organic and pesticides) were evaluated as well as normal operating conditions. The pesticide shock load condition was evaluated since the North Treatment Plant frequently receives variable amounts of pesticide type compounds from one industry.

The results obtained to this point are preliminary in nature; however, they do tend to confirm several aspects of the field studies. Detail comparisons are difficult to make at this time due to the different waste materials and difference in size of units. Directionally, the laboratory studies tend to confirm the following observations made in the field studies.

1. Oxygen uptake rate. The oxygen uptake rate in the aeration basin in both studies was below 50 mg/l/hr. Analysis of data from respirometor studies performed with biofilm removed from the aeration basin and the wheels indicated a stabilization phenomenon as opposed to a rapid growth condition of a fluid bed system. This substantially lower uptake rate represents a significant savings in operational costs over a fluid bed system. Supplemental aeration in the first contactor was necessary in the laboratory unit to prevent anaerobic conditions from developing on the first contactor. The recycling of high concentrations of settled MLSS resulted in a high uptake in the first contactor.

Figure 5

ACTIVATED ROTATING
BIOLOGICAL CONTACTOR SCHEMATIC



2. Stability Under Variable Loads. Using the synthetic feed, the organic load on the system was varied to produce a F/M ratio of 0.2 up to 5.0 based on COD into the system and MLSS in the aeration basin. The system consistantly achieved better than a 90% organic removal with F/M ratios greater than 0.8 while at lower ratios, the effluent deteriorated markedly as shown in Figure 6.
3. Resistance to toxic loads. During the field studies, the Maxson plant and the ABF pilot plant both received a shock load of a phosphate based pesticide. The fluid bed system exhibited a decrease in performance efficiency as was expected; however, the performance of the pilot plant was not affected. To evaluate further this phenomena under controlled conditions, a series of shock load studies using chloro-carbon intermediates from the manufacture of endrin was performed with the laboratory (3). Not only were the pesticide-type compounds absorbed onto the biofilm and thus removed from the liquid phase but biodegradation took place due to the hetrogenous growth in the biofilm. Two destinctly different gram negative bacilli and various yeasts were identified through enrichment culture studies using the chloro-carbon compounds as the sole carbon source. Neither pure cultures isolated from the primary enrichment media nor various reconstituted mixed cultures would use the chlorinated carbon compounds as a carbon source. However, four successive transfers of mixed cultures from the primary enrichment media resulted in heavy growth and chlorinated compound breakdown.

IV. SUMMARY

The results to date, both laboratory and field pilot plant, indicate that the activated biological contactor (ABC) concept offers several advantages over fluid bed systems in wastewater treatment. Considerable energy savings appear possible due to the lower oxygen uptake rate of the biomass. Although the oxygen uptake rate of the ABF tower bottoms was relatively high ($\sim$200 mg/l/hr.), the rate in the aeration basin following the tower was less than 50 mgl/hr. All of the reactors (contactors and aeration basin) in the laboratory study exhibit uptake rates less than 50 mg/l/hr.

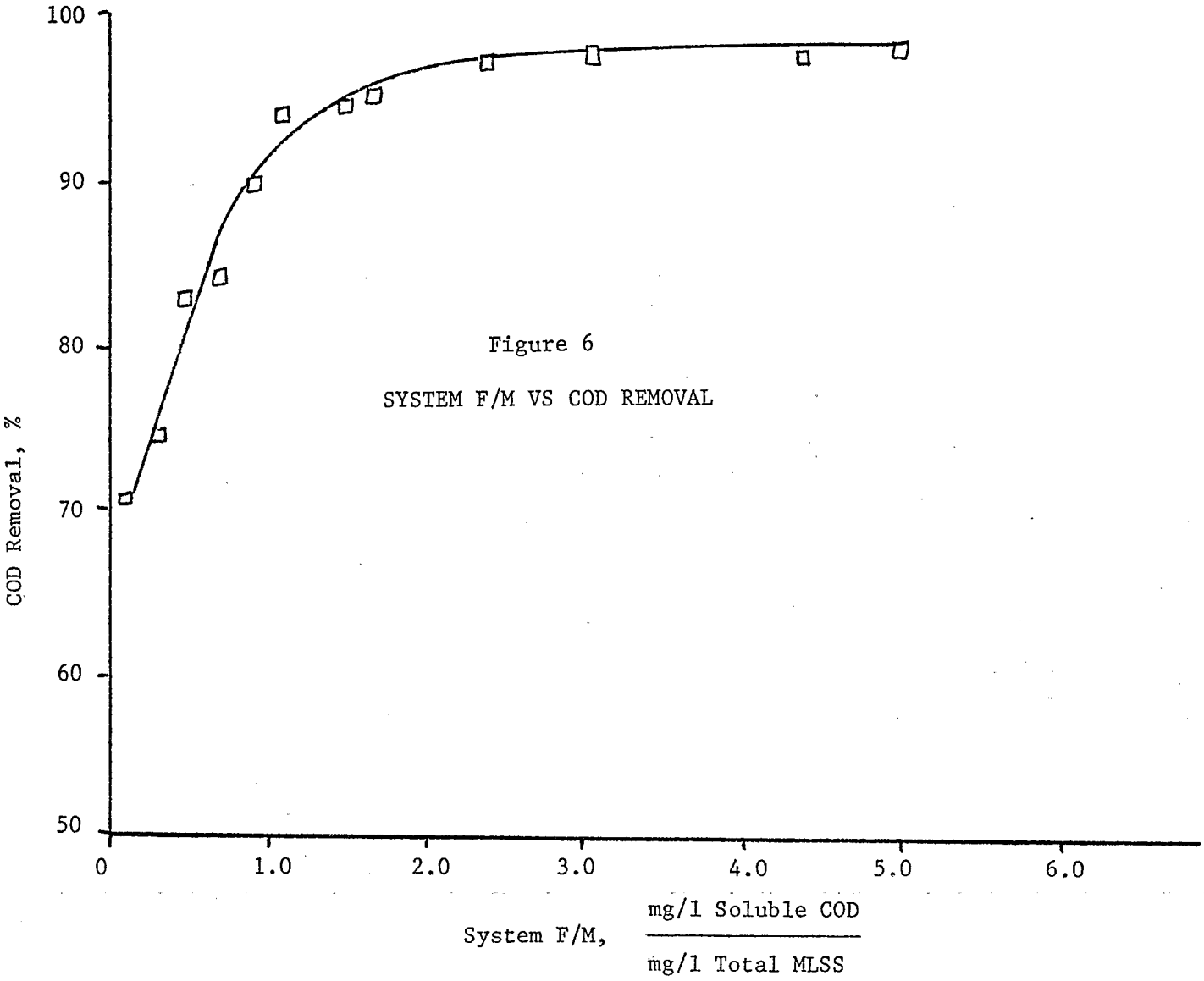


Figure 6

SYSTEM F/M VS COD REMOVAL

The hetrogenous biomass developed in the ABC system provides significant operational stability to shock loads of organic or toxic orgin. Several chlorinated compounds normally considered nonbiogradable and somewhat toxic were assimilated by the biofilm in the laboratory unit. Similar results were observed with a phosphate based pesticide shock load on the pilot plant system.

Work is presently continuing of Memphis State University to define further the mechanisms involved in the ABC. Additional data will be developed to explore the application of a mass transfer model to the system similar to that evaluated by Famularo (4).

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COMPARISON OF FIXED-FILM REACTORS WITH A MODIFIED SLUDGE BLANKET REACTOR

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INTRODUCTION

Over the last one hundred years, the anaerobic treatment process has been developed beginning with "Moura's Automatic Scavenger" and progressing to the conventional complete mixed and fixed-film anaerobic reactors (1). The conventional process is generally used for treatment of municipal sludges and other concentrated wastes. The advantage of this process is its simplicity in design and operation. The disadvantages are that a long hydraulic detention time is required for process efficiency and that reduced effluent quality results from a high concentration of suspended solids unless some means of effluent solids separation is provided.

To avoid some of these potential problems, the anaerobic contact process has been developed (2) where effluent suspended solids are settled and recycled back into the reactor. This leads to a longer solids retention time and therefore permits a significant volume reduction for a given treatment efficiency.

To be able to treat relatively dilute organic wastes, care must be taken to obtain a sufficiently long microorganism retention time within the system. Several alternatives have been proposed, such as the "anaerobic filter" by Young and McCarty (3), the "anaerobic attached-film expanded bed"

reactor by Switzenbaum and Jewell (4), the "upflow anaerobic sludge blanket" reactor by Lettinga (5), and the "anaerobic rotating biological disc" reactor.

Excellent results have been achieved with the "anaerobic filter" in laboratory investigations (3,6-7). The advantage of the system is its high reliability combined with high loading capacity and efficiency. However, the reactor requires filling material which influences the economics of the process. Furthermore, clogging problems may arise with the filter which can influence its reliability if run over a long period of time.

The "anaerobic attached-film expanded bed" reactor has the advantage of being relatively free of clogging, as the waste passes in an upward direction through a bed of suspended media to which the bacteria attach. However, the disadvantage is the rate of recycling generally required to keep the media in suspension.

The "upflow anaerobic sludge blanket" reactor is a modified version of the contact process and is based on an upward movement of the waste through a dense blanket of anaerobic sludge. This provides a greater surface area between the gas and the liquid which is advantageous in keeping the floating solids from clogging gas ports. Nevertheless, a large risk with this reactor is the possibility of further bed expansion and excessive loss of microorganisms to the effluent. The sludge blanket process also requires special granular sludge, which is difficult to develop.

The "anaerobic rotating disc" reactor has also been proposed for anaerobic treatment of wastewater (8). In this study, it proved to be a reliable and stable operating reactor with little potential for clogging while providing a high void volume. Its major disadvantage is high capital cost due to its relatively complicated construction.

Perhaps the major obstacles to wider application of the anaerobic process for industrial waste treatment are the relative difficulty in operation and the absence of a simple and cost-effective design.

The objective of this paper is to present a new process termed the "anaerobic baffled reactor" which is simple in form and may offer an economical solution to the treatment of intermediate and low strength industrial wastewater. Furthermore, a unified model for the mathematical description of fixed-film reactors and sludge blanket reactors will be presented.

BAFFLED REACTOR CONCEPT

The baffled reactor (BR) for anaerobic treatment is essentially a series of upflow sludge blanket reactors, but because of its unique characteristics, it requires no special granular growth of bacteria which is difficult to obtain. The baffled reactor's construction allows a high void volume and, therefore, clogging problems are essentially eliminated. A schematic diagram of an anaerobic baffled reactor is shown in Figure 1. This process evolved from initial studies with an "anaerobic rotating biological disc" reactor from which it was found that no rotation of the discs was necessary to obtain reliable reactor performance.

MATERIALS AND METHODS

Reactors. A comparison was made between the performance of three laboratory-scale anaerobic reactors: the anaerobic filter, the rotating biological contactor, and the baffled reactor. All reactors were constructed from plexiglas. The anaerobic baffled reactor (Figure 1) was 19.3 cm in length and had a total volume of 1040 cm^3 and an effective volume of 680 cm^3. The liquid passed horizontally and around baffles which served to maintain micoorganisms within the reactor. Sample ports were placed at 2.5 cm intervals along the reactor with an additional tap near the effluent port. The baffled reactor effluent was passed through an inverted siphon to separate the gas from the liquid. A low speed peristaltic pump fed the reactor. After the comparitive reactor evaluation, the anaerobic baffled reactor was scaled-up to a 6.3 liter liquid volume (Figure 2). Several modifications were made. The downflow chambers were narrowed, thus widening the upflow chambers where most of the cell mass had been found to collect. The lower edges of the baffles were slanted to route the flow to the center of the upflow chamber to achieve greater mixing of feed and solids. The number of ports was increased to facilitate sampling and wasting of solids should clogging occur.

The anaerobic filter (Figure 3) consisted of a bed of stones with average diameter of 12 mm through which the liquor was passed continuously in an upward direction. The filter had a porosity of 0.42 and a liquid void volume of 400 ml.

The anaerobic rotating biological contactor (Figure 4) consisted of circular plates connected to a slowly rotating horizontal shaft. Organisms attached to the surfaces of the

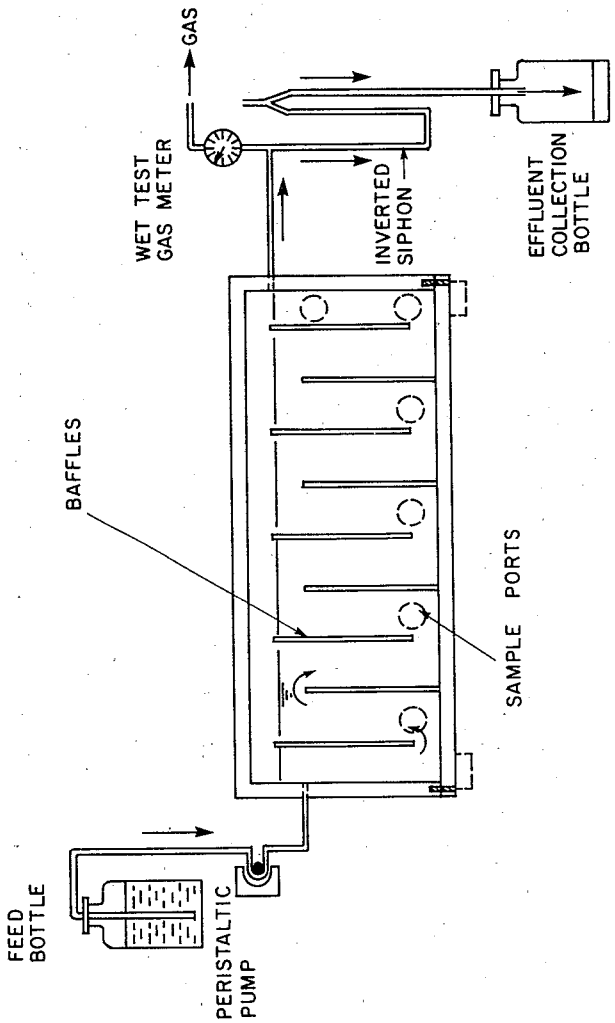

Figure 1. Schematic Diagram of Anaerobic Baffled Reactor System.

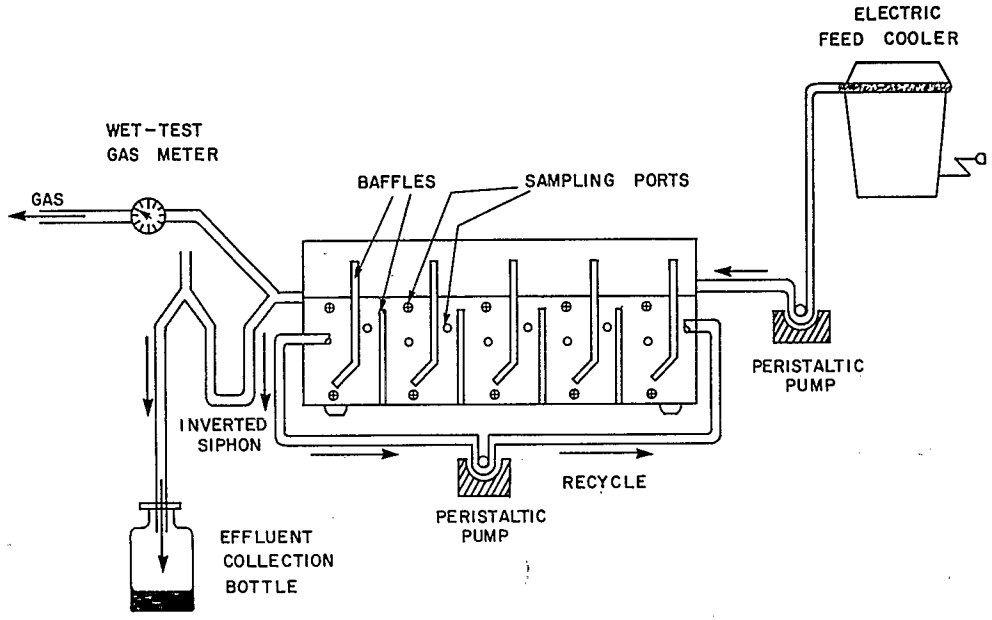

Figure 2. Schematic Diagram of Modified Anaerobic Baffled Reactor System.

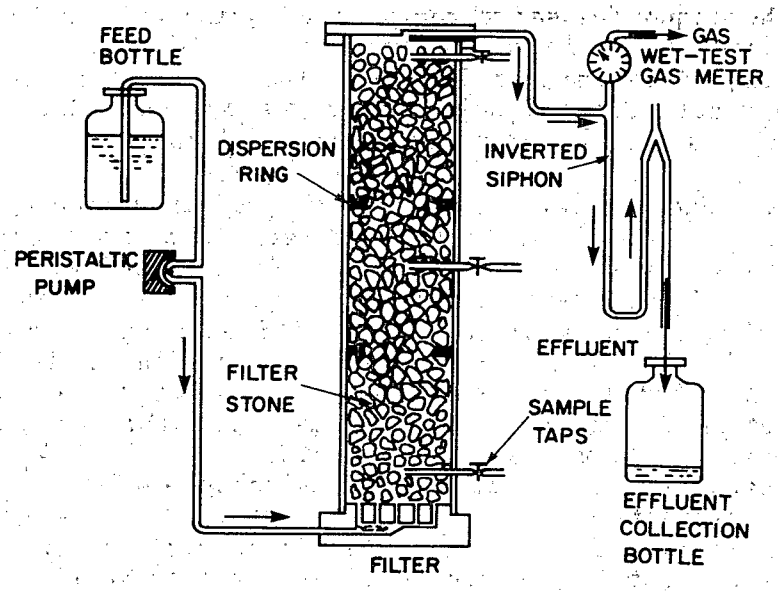

Figure 3. Schematic Illustration of Anaerobic Filter System.

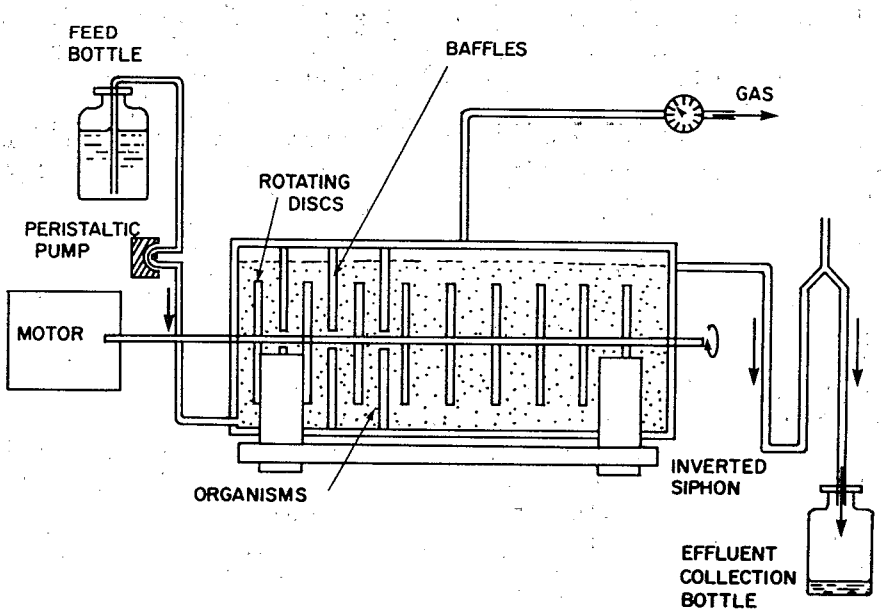

Figure 4. Schematic Illustration of Anaerobic Rotating Biological Contactor System.

plates. The liquor passed through in a horizontal direction and contacted the microorganisms on the plates. The reactor's liquid volume was 700 ml. The gas outlet was on top of the reactor; the liquid outlet was on the side. The outlet level was controlled by an inverted siphon.

Experimental Procedures. A complex protein-carbohydrate mixture was selected as the substrate for the study. Nutrient broth, an almost pure protein mixture, and glucose, a pure carbohydrate, were combined in equal chemical oxygen demand (COD) quantities to make up the feed with tapwater. Sufficient nitrogen and phosphorous were available in the nutrient broth for anaerobic growth. A sodium bicarbonate buffer solution was sterilized separately and then added to the feed solution to maintain the pH between 6.7 and 7.4.

The reactors were seeded with anaerobic sludge from a municipal treatment plant, and were operated at a constant temperature of $35^{o}C \pm 0.5^{o}C$ in a walk-in controlled-temperature chamber. At the same hour each day, gas production was recorded at atmospheric pressure with a wet-test meter (Precision Scientific Co., Model 63115).

Twice a week, routine laboratory measurements of samples withdrawn from the various levels of the reactors included COD, pH and total volatile acids. Gas composition was determined weekly by gas chromatography. Effluent alkalinity was occasionally monitored. Standard analyses were carried out according to Standard Methods (9).

Detailed results are presented first for the baffled reactor, and then its performance is compared with that of the other reactors.

RESULTS FOR BAFFLED REACTOR

Start-up and Loading. Initially, the organic loading of the baffled reactor was kept at 2 kg $COD/m^3 \cdot d$ and then was increased gradually up to a loading of 20 kg $COD/m^3 \cdot d$. Reactor performance was constantly monitored. Thereafter, the organic loading was stepwise reduced back to 5 kg $COD/m^3 \cdot d$. In a third phase, the hydraulic loading was varied, while maintaining the organic loading at a constant level of 5 kg $COD/m^3 \cdot d$. The loading scheme, together with the influent COD concentrations are shown in Figure 5.

Treatment Efficiencies. While increasing loading between 3 and 7 kg $COD/m^3 \cdot d$, the COD removal was essentially constant at about 78%, and decreased to 55% at a loading of 20 kg

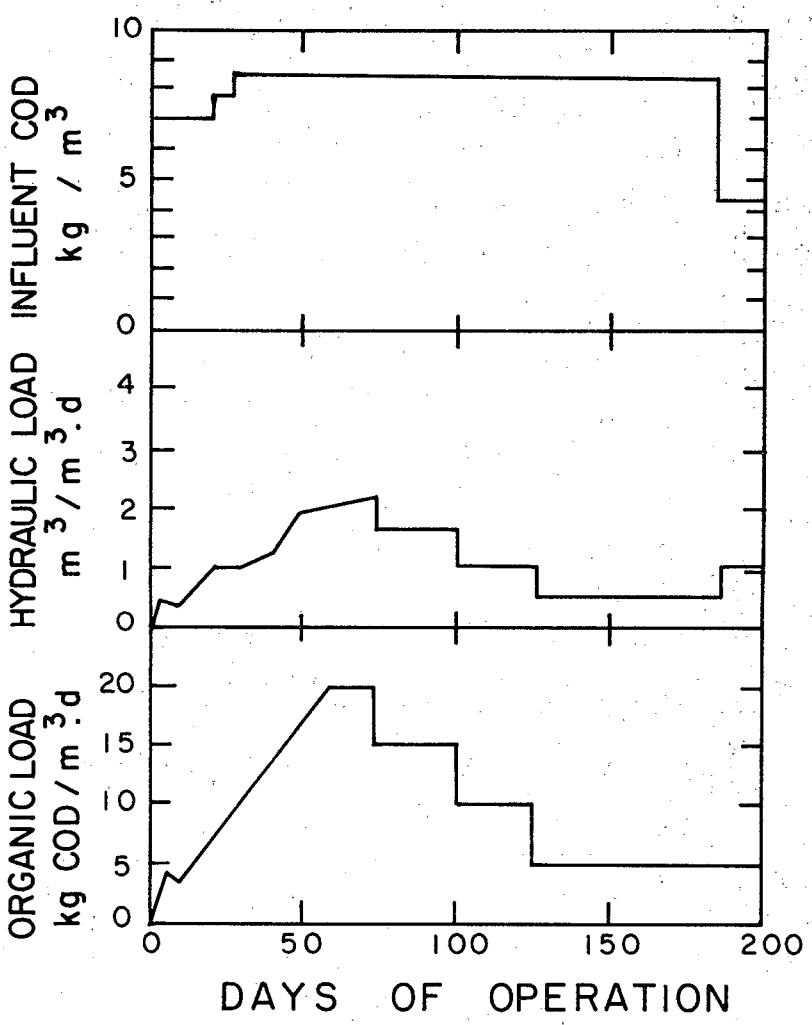

Figure 5. Loading Schedule for Anaerobic Baffled Reactor.

COD/$m^3 \cdot d$ (Figure 6). The COD removal was linear up to an organic loading of 12 kg COD/$m^3 \cdot d$. The maximum attainable COD removal rate with this small reactor was 10 kg COD/$m^3 \cdot d$ for organic loadings above 16 kg COD/$m^3 \cdot d$.

The decreasing loading range showed a significant deviation from the increasing loading range operation. It appeared that this was caused by the formation of gelatinous growth of bacteria at the head end of the reactor which caused short-circuiting of substrate through channels in the sludge blanket.

Gas Production Rates and Methane Content of Gas. The gas production rate (Figure 7) increased linearly with organic loading up to 10 kg COD/$m^3 \cdot d$. Above that, there was a slight downward deviation from the linear increase. The percentage of methane in the gas was 70% up to a loading of 7 kg COD/$m^3 \cdot d$ and methane content then decreased to 50% at a loading of 20 kg COD/$m^3 \cdot d$. The performance during the decreasing loading range showed only a small deviation from the increasing loading range for the gas production rate, whereas this difference is more evident for the methane percentage. Overall, the gas production rate and the methane content of gas coincide well with the COD removal rate and the treatment efficiency.

Modified Baffled Reactor. In order to improve the performance of the baffled reactor, a reduction in the gelatinous growth of bacteria that occurred in the head end of the reactor was needed. Also, high volatile acid concentrations occurring with high substrate concentrations had to be controlled to minimize buffer additions. Recirculation of effluent to dilute the influent waste concentration to about 5 to 10 kg/m^3 significantly reduced both of these problems. This is shown by the intitial results with the enlarged and modified baffled reactor (Table I). The treatment efficiencies and gas production rates are higher than with the smaller baffled reactor without recycle. Still higher loadings should be possible.

COMPARITIVE EVALUATION OF REACTOR PERFORMANCE

The three different types of laboratory-scale, high-rate reactors were investigated and their performance is compared in Table II. All data were taken at steady state, and the loadings and rates are based on void volume. Two of the reactors, the anaerobic filter (AF) and the anaerobic

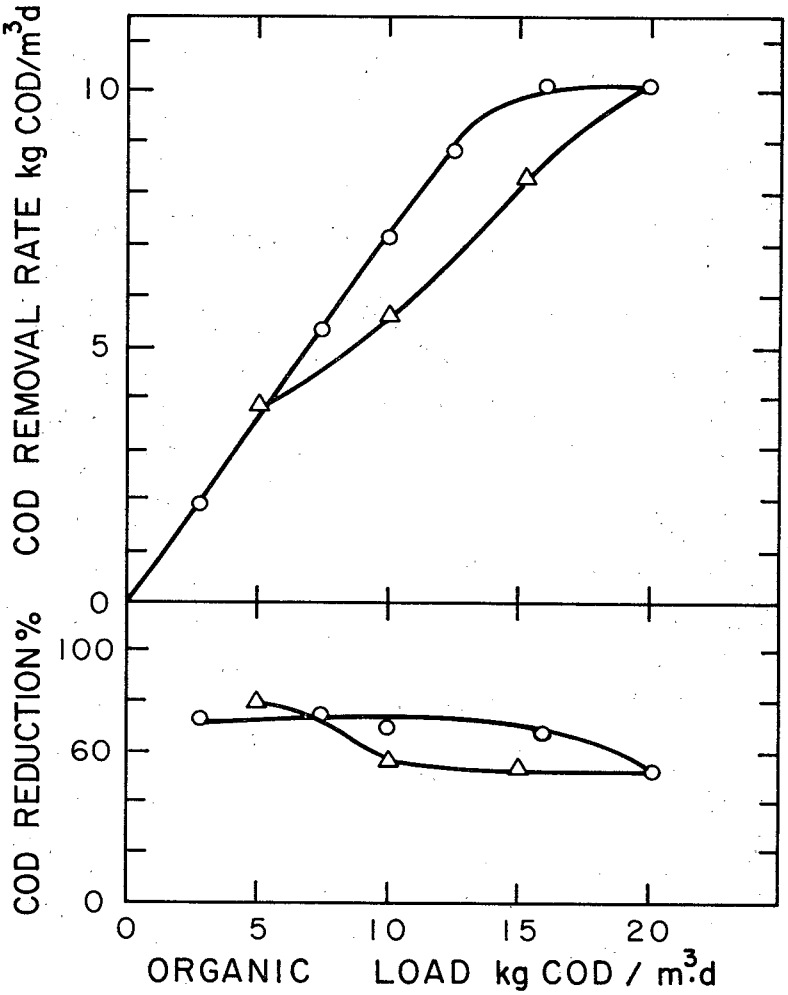

Figure 6. Baffled Reactor COD Removal Rate and Treatment Efficiency, Increasing Loading Range (Circles), Decrerasing Loading Range (Triangles), and Influent Substrate Concentration was 8.6 kg COD/m^3.

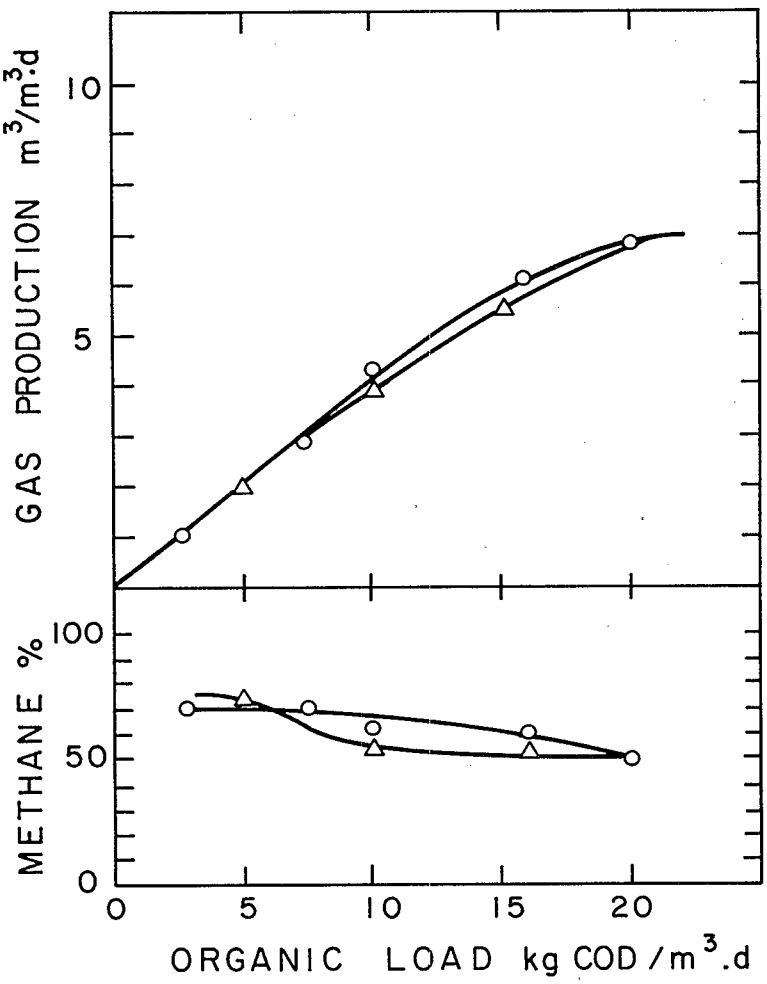

Figure 7. Baffled Reactor Gas Production Rate and Methane Percentage of Gas. Increasing Loading Range (Circles), Decreasing Loading Range (Triangles), and Influent Substrate Concentration was 8.6 kg COD/m^3.

Table I. Initial Results With Modified Baffled Reactor

Data Set	1	2	3	4
Influent COD Concentration, kg/m^3	7.3	7.6	8.1	8.3
Hydraulic Loading, $m^3/m^3 \cdot day$	0.5	1.1	1.1	1.3
Recycle Ratio: Qr/Q, m^3/m^3	0.0	0.4	2.3	2.0
Organic Loading, kg $COD/m^3 \cdot day$	3.5	8.3	9.0	10.6
COD Removal Efficiency, %	90	82	78	91
Gas Production Rate, $m^3/m^3 \cdot day$	2.3	4.5	4.3	6.9
Percent Methane, %	70	56	56	53
Effluent Volatile Acids, kg/m^3	0.34	0.80	0.70	0.40

Table II. Comparison of Reactor Performance

Reactor type	BR	Mod BR	AF	ARBC
Influent COD Concentration, kg/m^3	7.1	7.6	8.0	8.0
Organic Loading, kg $COD/m^3 \cdot day$	7.1	8.3	8.0	8.0
Hydraulic Loading, $m^3/m^3 \cdot day$	1.0	1.1	1.0	1.0
Percent Efficiency, %	79	82	92	90
Methane Production Rate, $m^3/m^3 \cdot day$	2.0	2.5	2.6	2.7
Percent Methane, %	70	56	80	78
Effluent Volatile Acids, kg/m^3	0.8	0.8	0.4	0.5

rotating biological contactor (ARBC) are considered to be fixed-film type reactors. The third type, the baffled reactor (BR) and the modified baffled reactor (mod BR), as already described, may be considered as complete mixed sludge blanket reactors (10). However, as shown in Table II, their treatment behavior under identical conditions appeared to be similar. For organic loadings between 7 and 8 kg $COD/m^3 \cdot d$, the treatment efficiencies varied between 80 and 90% and the methane production rates ranged between 2.0 and 2.7 $m^3/m^3 \cdot d$.

This finding was confirmed in a study by Frostell (11), who compared an AF system with a sludge blanket reactor. Over an organic loading range of 2 to 11.5 kg COD/m^3·d and a hydraulic loading range of 0.3 to 1.2 m^3/m^3·d, the behavior of the two reactors proved to be essentially identical, providing a good solids retention was achieved. This similarity in behavior led to an attempt to model two reactor types on a unified basis.

MODELING OF BAFFLED REACTOR PERFORMANCE

A fixed-film model was used to evaluate the hypothesis that the baffled reactor performance can be modeled as a fixed-film reactor. This evaluation is based upon the fixed-film model of Williamson and McCarty (12) as further evaluated by Rittmann and McCarty (13). The model provides a closely approximate, explicit solution for the flux of a limiting substrate into a "deep" fixed bacterial film. The model incorporates concepts of liquid-layer mass transport, Monod kinetics, and molecular diffusion. The derived equation provides an explicit solution for the flux into a bacterial film. The model has been applied with reasonable success to the AF (12). The reason this model was thought reasonable for the baffled reactor is that the sludge particles within the sludge blanket may be considered as fluidized spheres with surface area through which the solute must diffuse for bacteria consumption. The reactor may be considered as a sequence of five separate chambers connected in a series as illustrated in Figure 8.

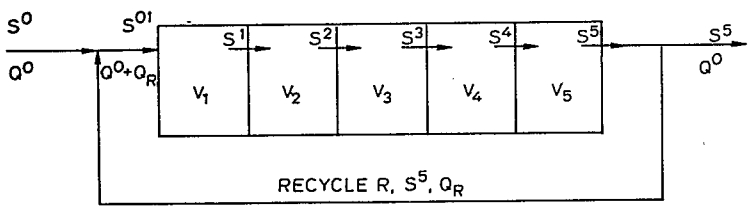

Figure 8. Flow Diagram of a Model of a Baffled Reactor.

For a complete-mix reactor with specific flow rate $Q(T^{-1})$ and specific surface area a (surface area per unit reactor volume, L^{-1}), a mass balance on substrate gives

$$\frac{dS}{dt} = -aCS^q + QS^o - QS \qquad (1)$$

where S is the bulk-liquid substrate concentration (ML^{-3}), S^o the influent concentration, C the variable-order reaction coefficient and q the variable-order reaction order. For the steady-state case, Eq. 1 can be solved algebraically for S:

$$S = S^o - (\frac{a}{Q})CS^q \qquad (2)$$

The model was applied by estimating the specific surface area a for each of the five chambers from profile-data of profile 1 and applying these values of a to predict the reactor behavior for other loadings (Profiles 2-4). This approach assumes a constant specific surface area a for each reactor chamber over the course of the study, which may not necessarily have been true. Equation (2) must be solved iteratively for S for each compartment of the reactor and due to the recycling, another iteration cycle needs to be done over the overall reactor, assuming a certain recycle substrate concentration. A mass balance of influent and recycle streams was used to determine the concentration in the stream entering the first reactor.

Figure 9 presents experimental results and model predictions for the kinetic coefficients and model parameters of Table III. The kinetic coefficients are based on the values of Lawrence and McCarty (15), and the model parameters on values from Williamson (16) and Williamson and Chung (17). Overall, the predictions were good, although the model as applied, resulted in a lower rate of removal for Profiles 2, 3 and 4 than actually found. This probably was partly due to use of a constant diffusion layer depth (L) for all cases. This depth would, however, decrease with increased gas mixing at higher loadings, leading to higher removal rates as noted. Also, the increased mixing at high loading rates would probably increase the area exposed to the substrate. Both of these corrections would lead to a better fit between predicted and measured performance.

A similar evaluation was performed assuming a series of completely-mixed dispersed growth reactors and using Monod kinetics. Values for active microorganism concentration were

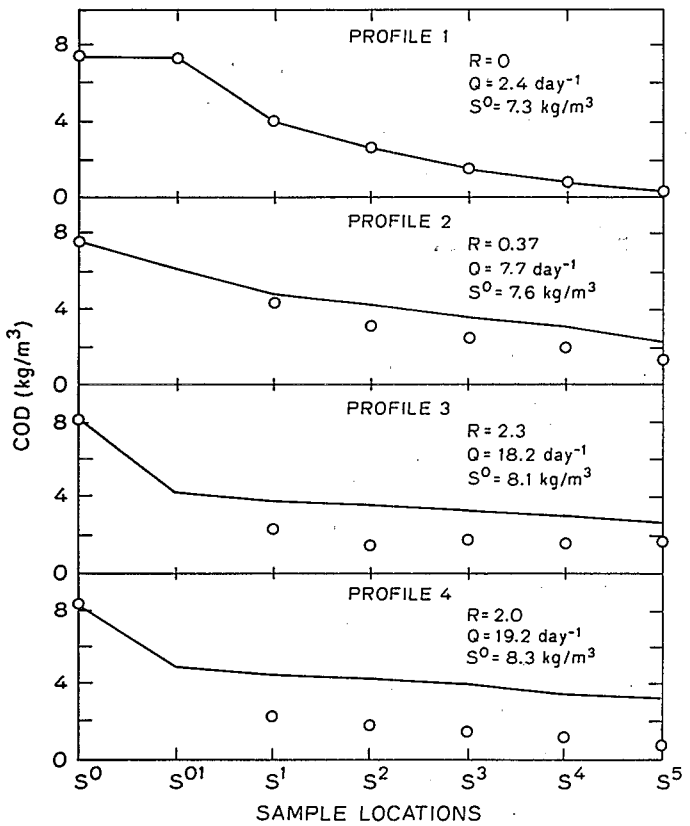

Figure 9. Comparison Between Steady-State Predictions Using Fixed-Film Model for Baffled Reactor and Experimental Data. T = 35°C, Kinetic Coefficients Used are Shown in Table III. Refer to Figure 8 for Flow Diagram of Baffled Reactor Model.

Table III. Kinetic Coefficients and Model Parameters

Process	Methanogenesis
Limiting Substrate	(acetate + proprionate) - COD
Temperature, ^{o}C	35
k, mg COD/mg VSS-day	8
K_s, mg/cm^3	0.2
X_f, mg VSS/cm^3 for chamber 1 to 5	10, 8, 8, 8, 5
L, cm	0.01
D_w, cm^2/day	0.8
D_f, cm^2/day	0.64

determined in each chamber for one loading case, and these values were used to predict substrate profiles for other loadings. The results as summarized in Figure 10 were poor. Such a model does not give realistic interpretations of the data as diffusional limitation in bringing substrate to bacteria is not considered.

SUMMARY AND CONCLUSIONS

An anaerobic sludge blanket process, termed the baffled reactor, has been developed which shows excellent promise for industrial wastewater treatment. It combines the advantages

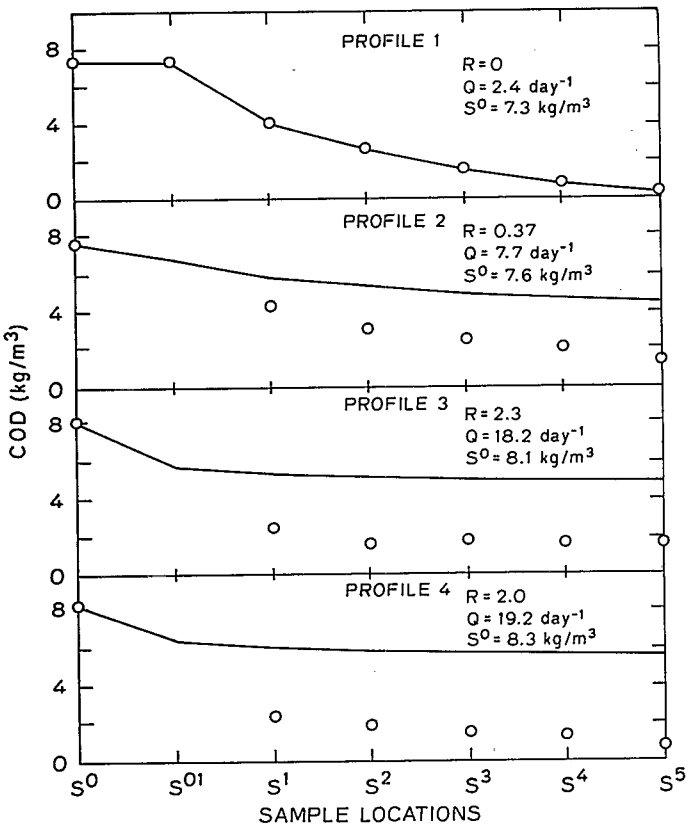

Figure 10. Comparison Between Steady-State Predictions Using Dispersed Growth Model for Baffled Reactor and Experimental Data. Refer to Figure 8 for Flow Diagram of Baffled Reactor Model.

of the anaerobic filter (3), which has a high stability and reliability due to attachment of the biological solids onto and between the filter media, and the upflow anaerobic sludge process (5) in which the microbial mass itself functions as the support medium for organism attachment, leading to a high void volume.

The baffled reactor's construction, however, avoids certain significant limitations of these other reactors. Specifically, the risk of clogging and the risk of sludge bed expansion with resulting high microbial losses have been minimized. The baffled reactor maintains a high void volume without the need of expensive and operationally work intensive gas collection systems or sludge separation systems. The over and under liquid flow reduces bacterial washout considerably, and does not require unusual settling properties for the microbial culture.

Although scale up factors are difficult to predict, the influence of gas stirring will be more important in large reactors due to the fact that gas is produced throughout the whole column height in the reactor. This should lead to greater evolution of gas per unit horizontal cross-sectional area, leading to more complete mixing in the upper portion of the chamber. It should also result in greater turbulence and resulting increase in mass transfer rates. These aspects should lead to better efficiencies in large scale reactors. However, other scale factors may decrease these advantages. Thus, large scale experiments are now needed for better evaluation of the baffled reactor patented.

The biofilm model appears to be generally applicable to sludge blanket reactors. This suggest a unified approach can be used in modeling several of the high rate anaerobic reactors. This aspect is of importance as the model is able to predict the performance from fundamentals of bacterial kinetics and mass transport. Although this preliminary work seems promising, future research is needed to confirm these observations and to better include effects of turbulence on mass transfer rates to biofilms.

ACKNOWLEDGEMENTS

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SYMBOLS

The following symbols are used in this paper:

a = specific surface area per unit reactor volume (L^{-1});
C = variable-order reaction coefficient (-);
D_w = molecular diffusivity in bulk liquid (L^2T^{-1});
D_f = molecular diffusivity in biofilm (L^2T^{-1});
k = maximum specific rate of substrate utilization ($M_sM_x^{-1}T^{-1}$);
K_s = half-velocity coefficient (M_sL^{-3});
L = length of effective diffusion layer (L);
Q = specific flow rate, (T^{-1});
q = variable-order reaction order, (-);
R = recycle ratio, (-);
S = bulk-liquid substrate concentration, (ML^{-3});
S_o = influent substrate concentration, (ML^{-3});
t = time, (T);
X_f = biofilm density, (ML^{-3});
V_i = volume of individual reactor chamber, (L^3)

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PART XI: AEROBIC AND ANAEROBIC TREATMENT-SUBMERGED MEDIA REACTORS

TREATMENT OF HIGH-STRENGTH ORGANIC WASTES BY SUBMERGED MEDIA ANAEROBIC REACTORS STATE-OF-THE-ART REVIEW

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INTRODUCTION

The anaerobic filter is basically an oxygen-free, media-filled bed reactor. Anaerobes grow not only in the void spaces between the media but also on the entire surface of the media. The wastewater can be distributed from the top (stationary type reactor) or it can be fed across the bottom of the filter (suspended type reactor). The latter type, the upflow reactor, is more popular than the downflow reactor. However, both reactors have completely submerged filter media that is arranged in either a packed bed or fluidized bed (Figure 1).

If anaerobic filters are classified on the basis of flow pattern, there are two main types: plug flow and complete mix. When the wastewater passes through an anaerobic plug-flow filter reactor, the pH decreases initially as a result of acid fermentation, and then increases in the direction of the process flow, due to the biological removal of the generated fatty acids, formation of ammonia and reduction of sulfates. Since the acidic pH in the bottom

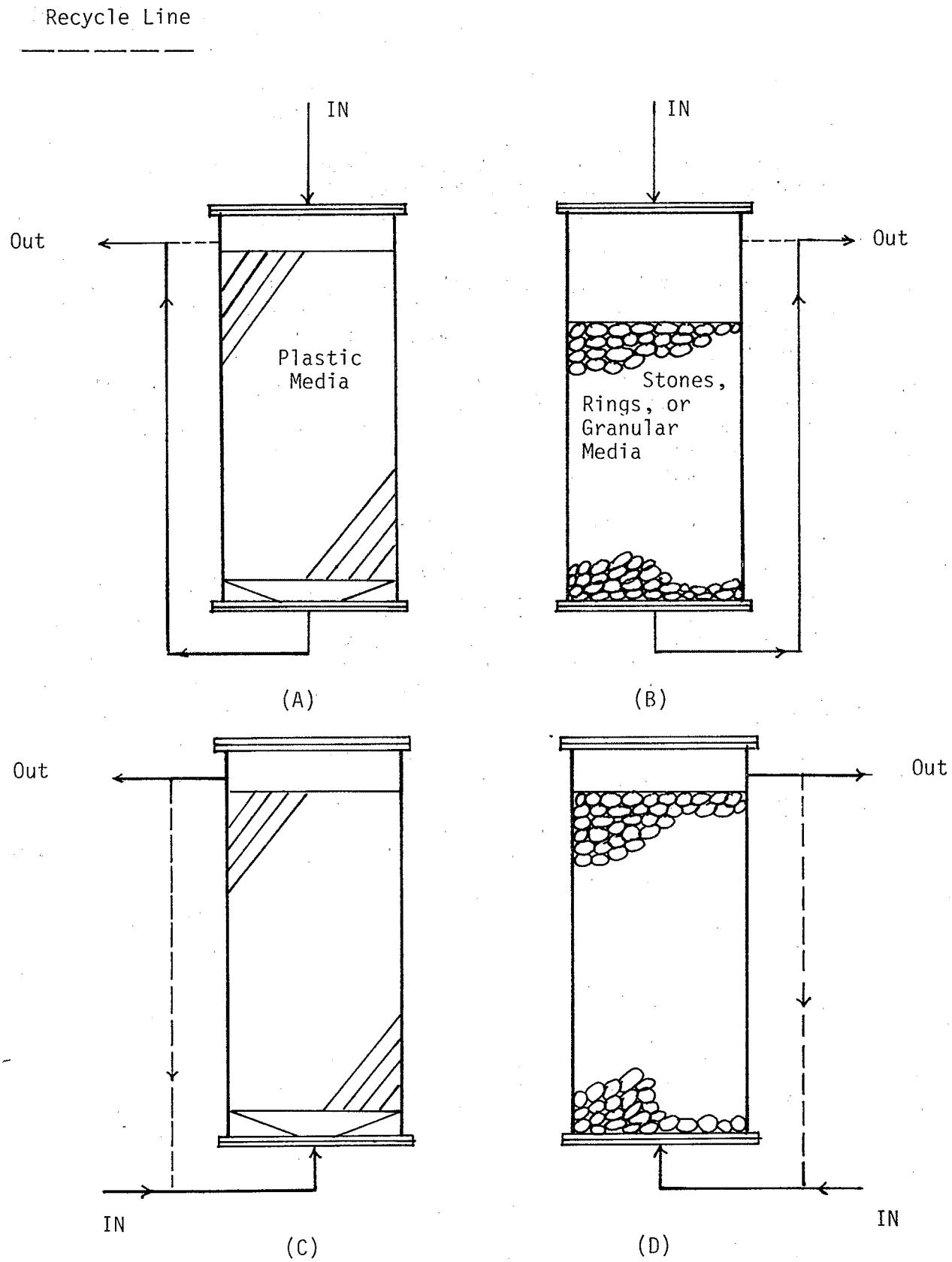

Figure 1. Schematic of Anaerobic Filters- (A) and (B) Downflow Reactors; (C) and (D) Upflow Reactors

section of the filter can potentially inhibit the methane-forming bacteria, substantial amounts of buffer solutions are added to the influent waste stream to prevent such pH decreases.

A completely mixed anaerobic filter would not experience the pH decrease observed in plug-flow units, since the mixing maintains a fairly uniform pH throughout the depth of the filter. The mixing of the filter is achieved by recirculating the effluent into the filter at a large recycle: feed ratio. This in turn would eliminate the need for adding costly buffer solutions. If the effluent has a sufficient bicarbonate buffer capacity, it is even able to neutralize feed solutions with an acidic pH.

The decomposition of waste water by the anaerobic filter has traditionally been considered to involve two stages. In the first stage, complex materials such as fats, proteins, and carbohydrates (COD) are hydrolyzed, fermented, and converted to simple organic acids and alcohols by facultative and anaerobic acid-forming bacteria. There is no waste stabilization during the first step because there is no methane production. Waste stabilization occurs in the second stage when the volatile organic acids are converted to carbon dioxide and methane by a special group of bacteria termed the methane formers. The methanogens are the most important group of bacteria because they carry out the final step in the overall process. They have slow growth rates and their metabolism is usually considered rate-limiting in the anaerobic stabilization of waste. However, the success of the process is dependent on the presence of both acid-producing and methane-producing bacteria, with the gas production stage responsible for stabilization of the organic materials.

Figure 2 represents the overall process of anaerobic digestion according to the traditional concept. While the distribution of carbon through the various types of intermediates may be correct, it is now recognized that the methane-forming bacteria do not utilize a variety of organic acids as substrate. Methane is formed from acetic acid, formic acid, CO_2 and H_2. Thus, the methane fermentation occurs only

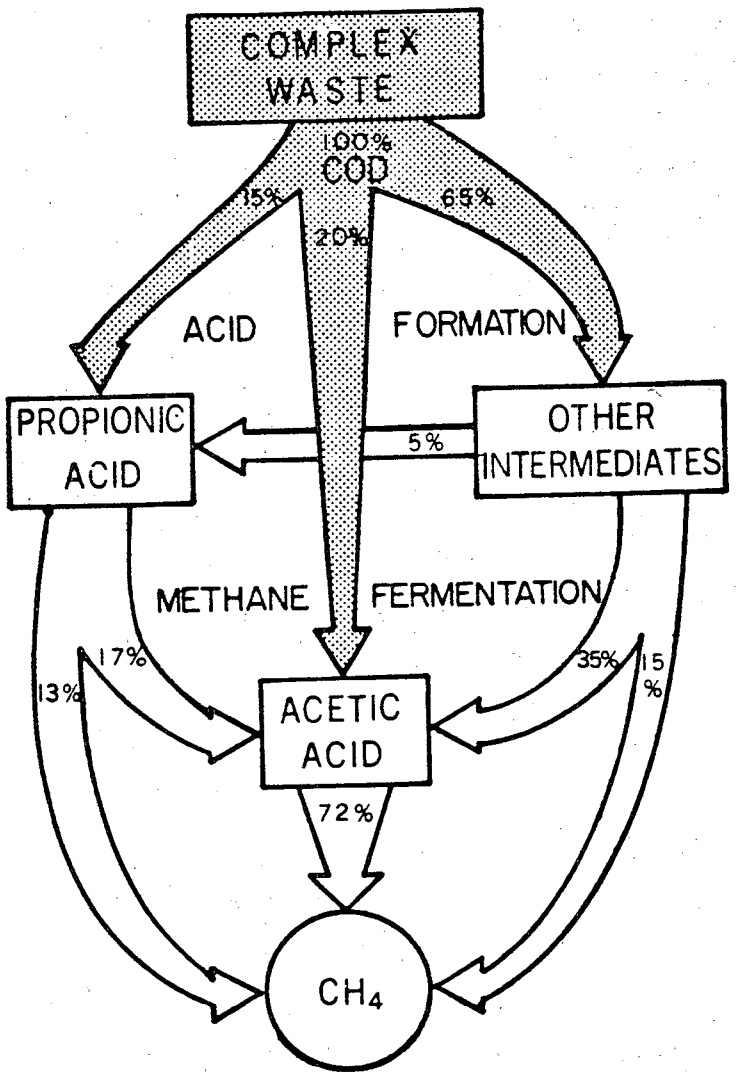

Figure 2. Metabolic Pathway In Methane Fermentation of Complex Organic Wastes [taken from McCarty (1)]

in the lower part of the figure and other organisms are responsible for the center portion, i.e., the conversion of various low-molecular weight acids and alcohols to the substrates useable by the methane-formers.

A major advantage of the anaerobic filter is its ability to produce methane gas. In general, the percentage of methane in the gas evolved from the filter is between 70 and 80%. Theoretically this gas could be used to heat the incoming wastewater or the reactor, thus increasing the efficiency of the filter and decreasing power requirements. The ability to use the methane gas in this way will depend on the quantity of gas produced. The volume of methane gas produced is approximately 4.3 to 8.0 ft^3 per lb COD stabilized.

The major factors influencing anaerobic filter performance include organic loading, hydraulic detention time, temperature, pH, alkalinity, wastewater characteristics, flow pattern, and type of filter media. These physical and chemical controlling factors generally have the same effects on the anaerobic filter as on the conventional suspended-growth anaerobic digestion system, except that the fixed-film anaerobic filter can be operated with higher organic loading, COD:N:P ratio, and metal concentrations.

Start-up of an anaerobic filter probably is the most difficult period of operation. Start-up times in experimental full-scale units have ranged from 10 to 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.

Analysis of the action of the process during start-up has indicated three factors of importance. First, the slow growth of anaerobic micro-organisms, especially at low waste concentrations and at temperatures below 30°C, does not permit rapid build-up of biological solids. Consequently, a large seed mass is needed for rapid start-up. Secondly, decreases in the buffering capacity of the waste, so that the pH drops below about 6.5 at any point within the filter for even short periods of time, increases the starting time significantly. A third factor affecting start-up time is related to the physical characteristics of the biological

suspended solids within the filter. During the early stages of operation a significant fraction of the biological solids remain finely dispersed throughout the liquid phase and a significant proportion washes out with the filter effluent. At some time after initial seeding , flocculation of the biological solids occurs in the filter and the solids washout rate decreases, thereby increasing both the rate of active biological solids accumulation and the rate of waste treatment.

The optimum method of seeding a filter is not known. Large seed volumes help to start the filter more rapidly by providing a large viable microbial population, and the large amounts of suspended solids help to promote the surface adhesion and flocculation which seem to be essential to good operation. However, using large seed volumes may also contribute significant amounts of volatile and non-volatile solids which tend to plug the filter and reduce its effectiveness for treating wastes.

CASE HISTORIES

Upflow anaerobic filters have been built in England since 1876 to purify sewage but the organic removal was mainly thought to be due to adsorption (2). Coulter, et al. (3) and Witherow, et al. (4) employed an anaerobic rock filter following an anaerobic sludge contactor. The combined processes produced a 65% reduction of BOD with most of it occurring in the first unit. Winneberger et al. (5) employed an anaerobic filter following a septic tank and noted a 65% BOD removal and a 70% suspended solids removal at a 5-day detention time. Research on anaerobic filters was published in 1968 by Young and McCarty (6). Loadings ranging from 26.5 lb COD/day/1000 cubic feet (0.424 kg/day/m^3* to 212 lb COD/day/1000 cubic feet (3.392 kg/day/m^3) were tested with theoretical detention times from 5.4 to 72 hours. COD removals ranged from 68% to 98%. Young and McCarty found that, at the same organic loading, the percentage of COD removal increased when the concentration of the influent COD increased.

*Kg COD/day/m^3 = 0.016 lb COD/day/1000 ft^3

Ham and Boyle (7) found that anaerobic treatment could effectively stabilize a raw leachate of approximately 10,000 mg/l COD with a detention time of 10 days and loading less than 32 lb COD/day/1000 cubic feet (0.512 kg/day/m^3). This system reduced COD by about 90%. A system with a 12.5-day detention time and loading of 13.0 lb COD/day/1000 cubic feet (0.208 kg/day/m^3) increased the COD removal efficiency to 93%. Foree and Reid (8) obtained a COD reduction of 96% for leachate with a COD of 12,900 mg/l. They obtained a higher degree of treatment at a loading of 80.2 lb COD/day/1000 cubic feet (1.283 kg/day/m^3). Chian and DeWalle (9) concluded that a high strength wastewater with an acidic pH can be successfully treated using a completely mixed anaerobic filter. Table 1 summarizes results of studies done with anaerobic filters.

Anaerobic treatment processes are very effective in removing heavy metals from waste streams by adsorption and precipitation. Digester studies have shown that heavy metals are present primarily in the solid phase as opposed to the aqueous phase. High metal removals result from the separation of the solids. A municipal digester study by Rudgal (32) found the influent copper concentration at 226 mg/l while the concentration in the effluent supernatant was 11 mg/l. The concentration in the sludge was as high as 500 mg/l. Chian et al. (33) investigated the anaerobic filter and obtained up to 95.5% metal removal. When these removals were calculated with respect to soluble concentrations, the percentage increased to 97.1%. This indicates that significant quantities of heavy metals are associated with the suspended solids leaving the filter. They concluded that their completely mixed anaerobic filter was effective in removing heavy metals. And as the metal concentrations increased in the effluent, the effectiveness of the filter increased. The metals are removed from the filter as a slurry in the bottom of the filter as they precipitate. They also indicated that with decreasing hydraulic detention time the metal removal percentage decreased while the metal content in the bottom slurry increased.

SOLIDS PRODUCTION AND EFFLUENT CHARACTERISTICS

A remarkable advantage of the anaerobic filter is its ability to retain active biological solids for long periods of time. There is a continual build-up of solids in the

Table 1 continued

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs.	% COD Removal	Effluent s.s. mg/l
Campbell 1973 (16)	Nonionic Detergent COD = 2,000 - 6,000 mg/l	25 - 75	48	55 - 83	20-950
El-Shafie & Bloodgood (17)	Metrecal COD = 11,000 mg/l	427	18	70 - 95	-
Dennis & Jennett (18)	Pharmaceutical Waste COD = 1250-16000 mg/l	14 - 220	12 - 48	94 - 98	10 - 60
Wilson & Timpany (19)	Dilute Waste Sulfite Liquor BOD_5= 1300 - 5300 mg/l	125 - 375	89 - 95	27 - 58% BOD Removal	-
Hovius et al. (20)	Synthetic Organic Alcohols, Aldehydes, Acids, Amine, Glycol, Phenol, COD = 2,000 - 20,000 mg/l	35 - 130	17 - 46	64 - 76	-

Table 1 Continued

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs	% COD Removal	Effluent s.s. mg/l
Foree et al. (21)	Brewery Press Liquor COD = 6,000 - 27,000 mg/l	50 - 400	15 - 330	30 - 97	110 - 500
Tadman (22)	Peptone-Glucose Brewery, Paper Mill Condesate Acetic & Glycolic Acid, TOC = 2,500 - 16,100 mg/l	62 - 450	12 - 48	73 - 94	-
Chian & Dewalle (23)	Landfill Leachate COD = 30,000 mg/l	-	648	97	-
Chian & Dewalle (9)	High Strength Acidic Leachate Containing Free Volatile Acids and Complex Carbohydrate like Organics COD = 12,900 mg/l	40 (recirculation ratio = 1:20)	1008	95	-

Table 1. Investigations of the Anaerobic Filter Performance

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs.	% COD Removal	Effluent s.s. mg/l
Young & McCarty (6)	Volatile Acid and Protein Carbohydrate COD=275-12000 mg/l	26.5 - 212	4.5 - 72	56 - 98	3 - 250
Plummer & Malina (10)	Food Processing Carbohydrate COD = 8,500 mg/l	100 - 640	13 - 83	30 - 86 55 - 86 soluble	450-1850
Clark & Speece (11)	Acetic Acid COD = 6,400 mg/l	370	12	30 - 80	-
Pailthorp,et al. (12)	Potato Processing Waste COD = 3,000 mg/l	33 - 145	13 - 59	41 - 79	250
Richter, et al. Taylor Dorstall (13 - 15)	Wheat Starch Waste COD = 2,000-6,000 mg/l	237	22	65 76% soluble COD removal	200-3360 (1330)

Table 1 Continued

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs	% COD Removal	Effluent s.s. mg/l
Foree and Reid (8)	Landfill Leachate COD = 12, 900 mg/l	80	240	96	40
Khan & Siddigi (24)	Non-Settleable Synthetic Liquid Waste COD = 2450 mg/l	225	8	81	-
Haug (25)	Waste Activated Sludge COD = 10,000 mg/l	-	48	76	-
Hakannson (26)	Whey (COD = 8,100 mg/l	119	-	97	-
Switzenbaum (27)	Sweet Whey Powder COD = 10,000 mg/l	1250	24 - 4	87%	-
Witt, et al. (28)	Acetic Acid Formic Acid 2- Butanol COD - 5,000 - 16,000 mg/l	250 - 380	24	87 - 90	-

Table 1. Continued

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs.	% COD Removal	Effluent s.s. mg/l
Witt (28)	Acetic Acid Acetaladehyde Glycol Vinyl Acetate COD = 7,000 - 10,000 mg/l	380 - 500	24	86 - 94	-
	Formic Acid Acetic Acid Methanol Formaldehyde COD = 17,000 - 24,000 mg/l	690 - 910	24	72 - 92	-
	Acrylic Acid Acrylate Esters COD = 79,000 - 85,000 mg/l	500 - 600	24	94 - 97	-
	Formic Acid Acetic Acid Methanol Formaldehyde COD = 17,000 - 24,000 mg/l	520	24	94.1	
	Dairy Waste COD = 24,000 mg/l	450 - 550	24	80 - 90	
Donnovan (29)	Heat Treatment Liquor COD = 11,540 mg/l	156 - 1250	14.4 - 96	50 - 80	-

Table 1 continued

Investigator	Waste Characteristics	COD loading lb/day/1000 ft^3	Retention Time hrs.	Removal	Effluent s.s. mg/l
Aurora, et al. (30)	Vegetable Tanning Effluent COD = 200 - 5,600 mg/l	6 - 204	72	90	-
Jewell, Morris & Switzenbaum (31)	Primary Settled Domestic Wastewater COD = 500 mg/l	87.5 - 288	5 - 8.3	85%	5

filter due to biological synthesis and no appreciable loss occurs until the filter becomes filled with highly concentrated biological solids. The filter can be operated for long periods of time before sludge wasting is needed because a low percentage of the COD removed is synthesized into biological solids. A material balance done by Chian and DeWalle (9) indicated that 93% of the COD removed could be accounted for by the methane gas formed.

Observation of the physical characteristics of the sludge within the filter indicates that the solids lie loosely in the interstitial spaces rather than becoming attached to the surface of the media. Plummer et al. (10) noted that solids in their filter units were not attached strongly to the media or to the sides of the unit.

It has been recognized that several factors determine the amount of solids leaving the anaerobic filter. Dennis and Jennett (18) observed that the solids concentration was mainly determined by the hydraulic detention time. Young and McCarty (6) observed a gradual accumulation of solids in the anaerobic filter, during which time the effluent suspended solids remained low. Only after the filter reached its maximum storage capacity would the effluent solids show an increase. When the influent waste contains solids, no net removal may be observed. In addition, it is noted that the porosity of the filter may have a large effect on the solids concentration which is to be expected since a lower void ratio will increase the collision frequency between the solids and the filter media.

The effluent characteristics cannot be categorized for all anaerobic filters. The properties of the effluent will be dependent on on-site conditions such as influent concentrations and loading rate. Most effluents contain a rather low concentration of suspended solids, a portion of which is readily settleable. All filters will have an effluent with a very low dissolved oxygen content. This is a disadvantage of the filter because the effluent cannot be discharged to the environment until the D.O. is raised to minimum requirements. The pH of most reactor effluents should range between 7.0 and 9.0.

BEHAVIOR UNDER ADVERSE ENVIRONMENTAL CONDITIONS

Anaerobic filters are much more resistant to variations in waste load and environmental factors such as pH and temperature than originally thought. Laboratory scale filters have shown rapid adjustment to four-fold surges in influent load.

There are four major types of transient loading or operating conditions that can affect filter performance: (a) variations in loading as a result of changes in flow rate or waste strength, (b) intermittent operation, (c) changes in pH, temperature, and waste composition, and (d) influx of organic toxins or heavy metals.

(a) Variations in Loading

Tests by a number of investigators (6,14,18) have shown that anaerobic filters can readily accept variations in load caused by changing either or both the flow rate or the waste strength, without being upset permanently. The following equation proposed by Young (34) suggests that changing the flow rate and waste strength simultaneously so that the organic load remains constant will not cause the effluent BOD to change.

$$S_e = \frac{K_1 S_o Q}{PV} = K_2 L \qquad (1)$$

where
S_e = effluent BOD concentration
S_o = influent BOD concentration
K_1 = proportionality constant
K_2 = proportionality constant
Q = flow rate
P = porosity of filter media
V = volume of reactor tank
L = organic loading to the filter

The equation also indicates that if the flow rate is held constant the "steady state" effluent BOD concentration will vary directly with changes in influent BOD. Data from El-Shafie and Bloodgood (17) and Dennis and Jennett (18) support this conclusion.

Short-term loading increases having a duration of one or two hydraulic detention times can be expected to produce only slight, short-term changes in effluent quality or gas production. Long-term changes, however, will cause the COD and volatile acid profiles, and the population dynamics and solids concentrations, to shift until a new "steady state" level of performance is reached. Four-fold instantaneous increases in loading have caused no permanent adverse effects on filter performance.

Chian and DeWalle (9) tested their complete mix anaerobic filter for its ability to withstand shock loads. When the detention time, based on feed stream flow alone, was reduced from 42 to 7 days for a 1-day period, only a small change in the pH was observed. The pH decreased from 7.2 to 6.9 when the detention time was reduced to 4.25 days for a 1-day period. The gas production did not show a corresponding increase, as high concentrations of organics were present in the effluent of the unit, which reduced the organic removal efficiency to 54%. Solids in the filter were resuspended at the higher flow rates which was indicated by the large differences in the values for the filtered and unfiltered COD and the increase in suspended solids. When the detention time was restored to 42 days, after the shock load, the effluent COD and suspended solids concentrations returned to values only slightly higher than those observed before the shock load. Based on these tests, it was concluded that the buffer capacity of the unit is sufficient to prevent large pH depressions at relatively short detention times. However, at detention times as short as three days, a large portion of the organics leave the unit in the effluent stream and the suspended solids experience an increase.

(b) Intermittent Operation

The second major type of operating condition that can affect filter performance is intermittent operation. The possibility has been tested by investigators (6,18) by stopping all flow and load to filters for several days, as might be used in practice for weekend operation, and there was essentially no loss in COD removal capacity or gas production efficiency upon restarting at full load. After

fourteen days of down-time without feed, the COD removal efficiency decreased to a greater extent, but full COD removal capacity and gas production were achieved after only three to four days of operation. Longer periods without feed might be expected to produce a low quality effluent for quite a long period of time after restoring waste load.

(c) Changes in pH, Temperature, and Waste Composition

Anaerobic filters, once "steady state" operation is achieved, become quite resistant to pH changes. Rapid recovery has been observed in filters exposed for a twelve hour period to pH levels as low as 5.4. Exposure to pH levels of 9.3 for as long as four days has caused only temporary loss of treatment efficiency. While gas production and COD removal were impaired at these extreme pH values, the filters recovered completely within one to two days after restoring pH to normal levels.

In general, anaerobic filters are expected to perform best at temperatures greater than 25° C. Filters have been used successfully to treat potato processing wastes at temperatures as low as 19° C, but too little information is available from which to draw significant conclusions about the effect of lower temperatures on filter performances.

Variations in waste composition are expected to produce little adverse response in anaerobic filters unless there is an associated influx of toxic materials. However, the composition of the waste significantly affects the solids produced in the system. Biological solids will accumulate much faster when treating a carbohydrate waste because synthesis of biological solids is greatest with carbohydrates, and at high loadings problems such as plugging or solids washout might be encountered. However, in studies to date, no filter has been reported to have become plugged beyond use.

(d) Organic Toxins or Heavy Metals

It has been generally assumed that anaerobic processes are unable to cope with waste streams containing toxicants and therefore are unsuitable for treatment of many wastewaters.

Toxicants do alter the kinetic parameters of methanogens and thus increase their generation time and decrease pollutant removal efficiency. However, these adverse effects can be offset by proper attention to solids retention time. Proper acclimation procedures can also increase the threshold concentration of toxicants which cause inhibition. The magnitude of the toxic effect generated by a substance can be reduced significantly if the concentration is increased slowly. This involves a process of acclimation which represents the adjustment of the biological population to the adverse effect of the toxin. The acclimation process in a mixed microbial population may involve any or all of three mechanisms: (1) mutation of one or more species in the population; (2) selection of the least sensitive species in the population; (3) alteration of the metabolism of one or more species to overcome the metabolic block produced by the toxic material. All these mechanisms may interact. In any case, resistance to a toxic substance often involves an increase in the concentration of the substance which can be tolerated rather than acquisition of total resistance to the substance at any level. When the concentration of a toxic substance is increased slowly, the microbial population can acquire increased resistance through all of the mechanisms available to it. However, if a large concentration of toxic material is introduced suddenly, the effects are quite different than when the same concentration is reached after an adequate series of acclimations because no time is allowed for any of the available mechanisms to operate, and most of the population will be destroyed. In evaluating data from toxicity studies for design purposes, the engineer should consider the test conditions used and whether toxic materials may be introduced into the waste stream to be treated as a slug dose of high concentration or as a constant component to which a population may become acclimated. Speece, et al. (35) showed that methanogenic bacteria could acclimate to toxicant concentrations that were 100 times greater than the concentrations which caused inhibition of unacclimated cultures. They found that continuous increments of nickel chloride could be added to the feed of a filter with no adverse effect on gas production. An increase from 200 to 400 mg/l decreased gas production but it resumed when nickel additions were stopped. They were able to acclimate the filter to a sulfide concentration of 1000 mg/l. Long term acclimation of

a filter to increasing levels of continuous sodium additions showed no adverse effect up to additions of 7500 mg/l as Na^+. Formaldehyde added to the feed of the anaerobic filter caused no inhibition of gas production up to 400 mg/l. Additionally they were able to acclimate the filter to 600 mg/l of acrylic acid and to acrolein at a concentration of 100 mg/l.

Parkin, et al. (36) found that cyanide and ammonia toxicity were fairly reversible while chloroform, formaldehyde and sulfide exhibited some irreversible toxicity. Nickel showed signs of irreversible toxicity depending on the concentration.

The early warning of possible metal toxicity is given by a gradual decrease in gas production and an increase of the effluent COD. This can be anticipated if strict attention is given to the influent wastewater so that the operator will know if inhibitory concentrations of metal ions are entering the filter.

KINETIC ASPECTS

Because of high cell mass concentration and immobilization of cells within the attached film, the substrate utilization rates per unit volume of the biological reactor are high; displacement of the culture composing the film by inactive cells in the influent to the reactor is less likely than displacement of a culture in dispersed/suspended growth, and a fixed-film reactor is less susceptible to upset by shock loadings of substrate and/or toxic metals.

As a result of these advantages inherent in the attached-growth wastewater treatment systems, many researchers have attempted to model growth and substrate removal by biofilms. Earlier studies described the removal kinetics in terms of substrate concentration existing in the bulk liquid phase. Recent investigations have generally described substrate removal kinetics in terms of simulated substrate concentrations in each layer of the biofilm. The response of the entire film is predicted by the sum of layers. The most complete biofilm models were developed by Williamson and McCarty (37) and Dewalle and Chian (38) on the basis of two competing mechanisms, diffusion

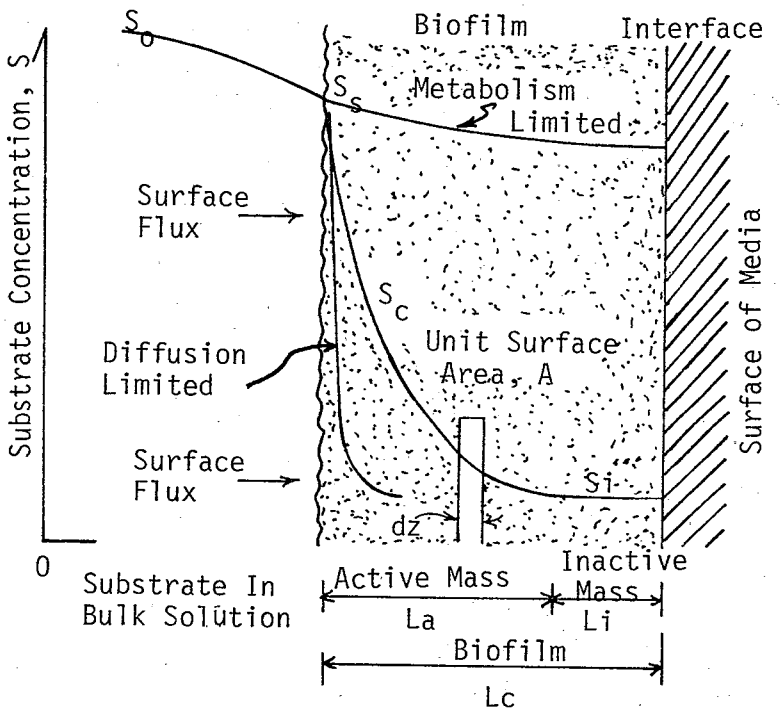

Figure 3. Substrate Concentration Profile Within A Biofilm [taken from Williamson & McCarty (37)]

and metabolism. Basically, all models considered mass transport from the bulk liquid to the biolayer by diffusion through a boundary layer of stagnant liquid covering the biolayer. (See Figure 3.) Williamson and McCarty's model is represented by a second-order, non-linear ordinary differential equation.

$$\frac{d^2S_c}{dz^2} = -\left(\frac{dS_c}{dt}\right)\frac{X}{D_c} = \left(\frac{qS_c}{K_s+S_c}\right)\frac{X}{D_c} \qquad (2)$$

The equation states that the second derivative of substrate concentration with respect to biofilm depth z is directly related to the utilization rate of the rate-limiting substrate ($-dS_c/dt$) and biomass concentration (X), but is inversely related to the diffusion coefficient (D_c) within a biofilm. Additionally, in Eq. 2 it is also assumed that the rate of utilization of substrate at any depth within the biofilm follows the Monod relationship; that is $-(dS_c/dt) = (qS_c)/(K_s + S_c)$, in which q is the maximum utilization rate of the rate-limiting substrate S_c and K_s is the Monod half-velocity coefficient.

This equation does not possess an explicit solution. However, it can be solved for the two limiting cases of the Monod equation. When the substrate concentration S_s (at $z = 0$) is much greater than the half-velocity concentration, K_s, Eq. 2 becomes a zero order kinetic equation:

$$\frac{d^2S_c}{dz^2} = \frac{qX}{D_c} \qquad (3)$$

and the biofilm substrate concentration (S_c) can be computed by using the equation given below:

$$S_c = S_s - \frac{qX}{D_c}\left(L_c z - \frac{z^2}{2}\right) \qquad (4)$$

But, when S_s is much less than K_s, Eq. 2 becomes a first order kinetic equation:

$$\frac{dS_c^2}{dz^2} = \frac{qXS_c}{D_c.K_s} \qquad (5)$$

Under the above condition, the relationship between the biofilm substrate concentration, S_c, and the other controlling parameters such as q, X, K_s, L_c, and z is defined as:

$$S_c = S_s \frac{\text{Cosh } [(qX/D_c K_s)^{\frac{1}{2}}(L_c - z)]}{\text{Cosh } (qX/D_c K_s)^{\frac{1}{2}} L_c} \qquad (6)$$

Obviously, the problem in using this biofilm model is the need to determine the thickness of the diffusional boundary layer (L_c) and the values of maximum substrate utilization rate (q) and diffusion coefficient (D_c). These are difficult to measure in a fixed-film biological system.

Because of these reasons, model modifications were suggested by DeWalle and Chian (38). According to Fick's law of molecular diffusion, the mass transfer rate ($\partial M/\partial t$) through a surface area A is proportional to the concentration gradient of the substrate at the interface:

$$\frac{\partial M}{\partial t} = -AD_c \frac{\partial S_c}{\partial z} \qquad (7)$$

By substituting the term $\partial S_c/\partial z$ in Eq. 7 into the integrated form of Eqs. 3 and 5 for a unit cross-sectional area and at $z = L_c$, DeWalle and Chian were able to define the rates of mass transfer as follows:

$$\frac{dM}{dt} = qXL_c \quad , \quad \text{if } S_s >> K_s \qquad (8)$$

and

$$\frac{dM}{dt} = S_b \sqrt{\frac{D_r qx}{K_s}} \quad , \quad \text{if } S_s << K_s \qquad (9)$$

where S_b is the substrate concentration in the bulk liquid. L_c in Eq. 8 can be approximated using the equation proposed by Pirt (39) and Saunders and Bazin (40) :

$$L_c = \sqrt{\frac{2 D_c \cdot S_b}{qX}} \qquad (10)$$

Eq. 8 indicates that the rate of mass transfer is independent of substrate concentration, but directly proportional to the thickness of the biolayer (L_c) and the concentration of attached biomass (X). If cell attachment is uniformly distributed, the rate of substrate removal is also proportional to the specific surface area of solid medium because Eq. 8 is derived from a unit cross-sectional area. More importantly, Eq. 8 further indicates that at very high substrate concentration the rate of substrate removal is highly dependent upon the specific surface area of solid medium within the system due to the zero order kinetics of the reaction. This is true in particular when submerged filters are employed for the treatment of high-strength organic wastes.

Eq. 9 states that the rate of mass transfer is independent of biofilm thickness, but directly proportional to the bulk substrate concentration (S_b) and the square root of the biomass concentration (X). Since for a given substrate, q, D_c, X, and K_s are not expected to vary greatly, Eqs. 7 and 8 can be reduced to:

$$\frac{1}{V}\frac{dM}{dt} = K_1 \frac{A}{V} S_b \qquad (11)$$

and

$$\frac{1}{V}\frac{dM}{dt} = K_2 \frac{A}{V} S_b \qquad (12)$$

where V is the reactor volume and A/V is the specific surface area in the reactor. K_1 and K_2 are coefficients based on zero and first order kinetics, respectively.

The effects of effluent characteristics, specific surface area and flow velocity on the substrate removal rate were discussed by the same investigators. They concluded that: (a) At low substrate concentrations the removal rate as predicted by Eq. 11 increased linearly with effluent concentration. The effluent concentrations have a finite value when the removal

rate approaches zero. On the contrary, at high substrate concentrations a satisfactory linear relationship was not obtained from Eq. 11 when the reciprocal substrate removal was plotted versus that of the substrate concentration. (b) An increase in K_2 was always observed as a result of increasing specific surface area although no definitive correlation between the rate of substrate removal and the value of the increasing specific surface area could be established. (c) The calculated K_2 (V/A) values appeared to have a straight relation with flow velocity when plotted on log-log paper. This means that the rate of substrate removal can be improved by minimizing the diameter of the submerged filter column or maximizing the height of the column in order to obtain higher velocities. Practically speaking, a tall column has one obvious disadvantage, i.e., a relatively larger volume in the last portion of the column is in contact with a low substrate concentration, which in turn tends to reduce the substrate removal rate.

Kinetic models are presently not supported by sufficient analytical data and it is necessary that more research be done in this area.

SUMMARY

The advantages inherent in the anaerobic filtration process suggest that it is worthy of consideration as a basis for full scale waste treatment facilities. The fixed film anaerobic process is well suited to handling large organic loads. High COD removals can be achieved, particularly with high strength wastes, without the high operating costs that are associated with other treatment operations which use aeration or physical-chemical methods. Anaerobic digesters also product a useable methane gas which could be used to heat the reactors. The methane production is almost certain to be well in excess of the necessary requirements for heating purposes and could possibly be used for additional heating of the treatment facility. Research has shown that the filter can successfully treat many types of waste efficiently. More quantitative data from field installations and laboratory pilot plants are needed to establish design criteria for this treatment process.

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