Use of the Rotating Biological Contactor for Appropriate Technology Wastewater Treatment

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ABSTRACT

The performance of a Rotating Biological Contactor (RBC) was studied at a loading rate of 0.1 gallon/ft.^{2}/day. The BOD_5 of domestic wastewater was reduced from 150 mg/ℓ to 3 mg/ℓ, a reduction of 98%. The total suspended solids were reduced from 73 mg/ℓ to 32 mg/ℓ and ammonia was completely oxidized to nitrate. The economics of wastewater treatment at this low loading rate will be favorable for applications which require maintenance free operation, or where operational expertise is unavailable.

A mathematical model for the RBC was also developed. The model includes material balances on both oxygen and substrate in the biofilm and bulk solution. The resultant set of non-linear, parabolic partial differential equations were solved using an implicit numerical technique similar to Crank-Nicolson. The model predictions were only 10% different than the experimental results. The model should provide a basis for future development.

TABLE OF CONTENTS

LIST OF FIGURES

I. INTRODUCTION

The Rotating Biological Contactor Process (RBC) has been used increasingly in Europe for the treatment of domestic wastewaters and for some industrial wastes. The process has found more recent application in the United States for similar types of waste particularly those from dairy and food processing industries. Some oil companies have given consideration in using them to treat wastes from refineries. Small communities which are located far away from major treatment plants have considered about using them.

The RBC consists of a series of discs attached to a common shaft. The discs are partially submerged in a trough of continuously flowing wastewater. As the discs rotate, a film of microorganisms growing on the discs consume oxygen from the air and substrate from the wastewater. In this way, organic materials (substrate) are removed from the wastewater.

The advantages claimed for RBC are: simplicity of maintenance and operation, low power consumption, no flies or objectionable odors, ability to withstand shock or toxic loads and desirable sludge settling properties.

In this paper, a dynamic model and results of an experimental investigation are presented. This model describes the removal of substrate using a material balance over the trough and liquid film, and diffusion of substrate and oxygen in the microbial film when submerged in wastewater and exposed to air. Since the diffusion equations are

nonlinear parabolic equations, numerical analysis was employed to solve them simultaneously. The predictions of the model are discussed and compared with data obtained from the UCLA pilot plant study.

This project was sponsored by the University of California Appropriate Technology Program. The RBC is suitable for Appropriate Technology because there exists a universal need for economical wastewater treatment in all areas of the United States. In many large cities, where a large nucleus of technical expertise and know-how exists, wastewater treatment has taken the form of sophisticated and energy intensive treatment plants which require few, but highly trained operators. The application of similar high technology to small countries has resulted in poor performance, due to a lack of technical knowledge and support. There exists a need for efficient wastewater treatment processes which do not require sophisticated operators and which conserve natural resources.

II. EXPERIMENTAL EQUIPMENT AND ANALYTICAL TECHNIQUES

Experimental Equipment:

In order to verify the mathematical model, data were collected from a pilot plant using domestic sewage as the substrate. In using sewage rather than synthetic substrate, the model and experimental results are more meaningful since the results are applicable to real situations in wastewater treatment plants.

The Rotating Biological Disc pilot plant was purchased from Autotrol Corporation in Milwaukee, Wisconsin in June, 1978. It consisted of a hemicylindrical tank made of fiberglass. The tank was divided into five stages. The last four stages were each 13 inches long and 11½ inches in radius with a volume of 9.25 gallons. The first stage was 9 inches long and was intended for temporary wastewater storage. A central steel shaft ran through the whole length of the tank (62 inches) and was used to support the polyurethane discs. For each stage there were 9 discs. Each disc had a diameter of 18-5/8 inches and were attached together in each stage to allow maximum surface area for a given volume. The total surface area for 36 discs was 250 ft^2, providing a volume to surface area ratio of 0.148 gallons/ft^2. The discs were 40% submerged and rotated at a constant speed of 7 RPM, providing a peripheral velocity of 34.1 ft/min.

Wastewater normally flows from stage to stage through one inch diameter holes in the baffles which seperate the stages. At the low flow rates used in this investigation, the one inch diameter hole

would permit excessive backmixing. It was necessary to plug the one inch diameter hole and connect the stages through U-shaped PVC fittings. A hydraulic gradient was set for these elbows so that substrate flowed in only one direction. The buckets initially designed for fixed flow rate from the small reservoir in the beginning of the tank were removed and substrate was fed through a tube, which was connected to a variable speed Master Flex pump. The substrate was pumped from a drum where primary settling took place. Fresh domestic sewage was fed into the drum daily. The experimental set up is shown in Figure 1.

Domestic wastewater was collected from the Westwood Boulevard sewer at the UCLA campus and was found to be typical for domestic wastewater with an exceptional high organic nitrogen content. Since organic nitrogen lowered pH when it was converted from ammonia to nitrate, sodium carbonate was used to make sure the effluent pH would not fall below the effluent standard. The pilot plant was operated inside the research laboratory so that the RBC could be maintained at room temperature. Room temperature was maintained between 13-30^{o}C throughout the experiment.

Analytical Techniques:

Different kinds of data were collected in this experiment. Ammonia, pH, and nitrate data were collected 5 days a week; BOD_5, COD, and TSS data were collected 2 days a week throughout the experiment. Nitrite and phosphate data were only collected twice and it was found to be sufficient.

For BOD_5, a dissolved oxygen (D.O.) meter made by Yellow Spring Instrument was used to measure the initial and final D.O. Before each

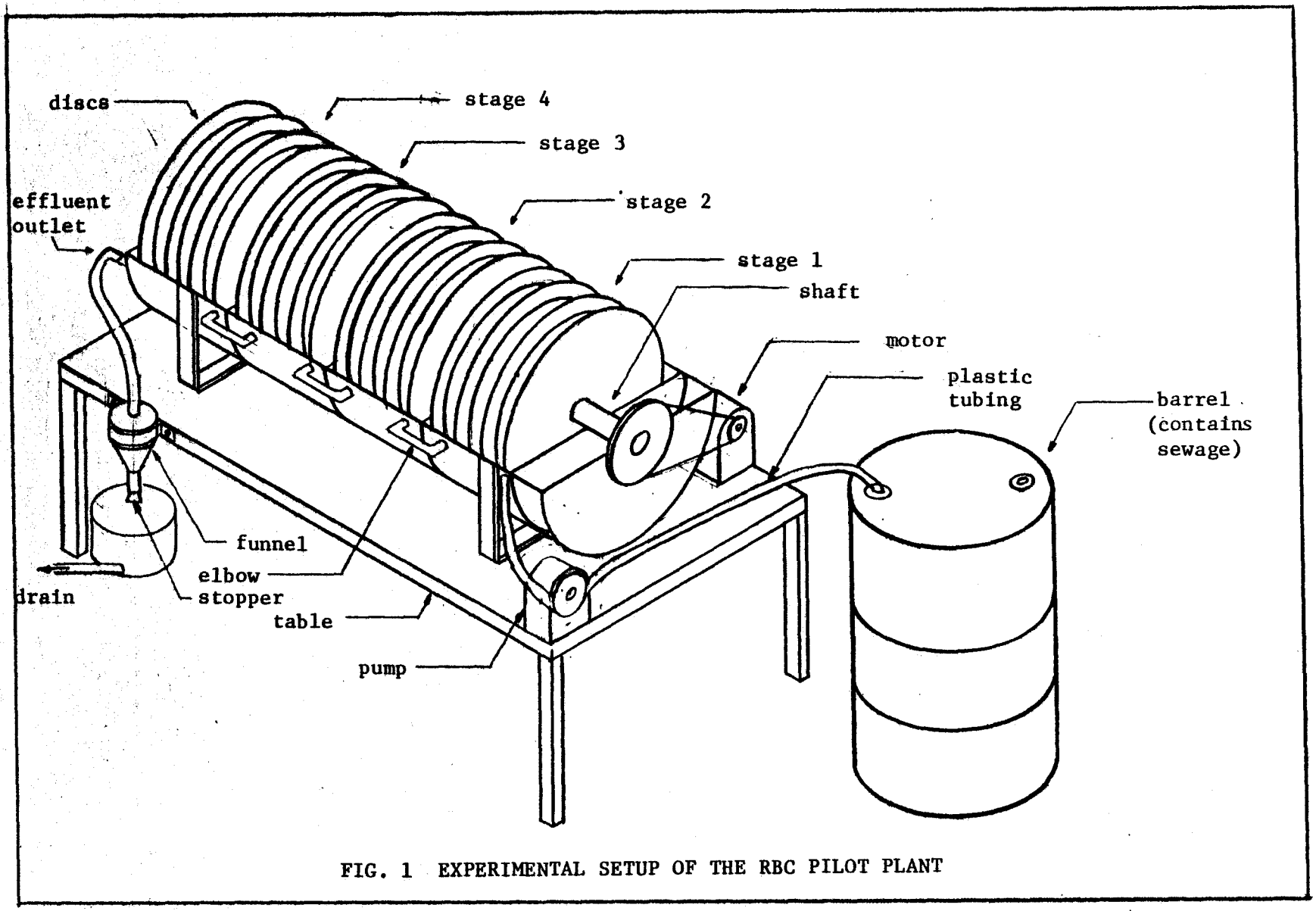

FIG. 1 EXPERIMENTAL SETUP OF THE RBC PILOT PLANT

measurement, the meter was calibrated by oxygen-saturated water whose D.O. was found by the Winkler method in the Standard Methods (Method 507). (1)

For pH, a model Corning 12, pH meter was used. For ammonia and nitrate, an Orion Research 407A meter with specific ion electrode was used. Before measurement, the meter was calibrated by standard solution made by the same company. For COD, the potasium dischromate method was used (Method 508), for TSS (Method 208A). For nitrite, the colorimetric method was used (Method 420). For phosphate, the stannous chloride method was used with initial persulfate digestion (Method 425C III and 425E).

III. REVIEW OF RELEVANT LITERATURE

Operational Data:

The Rotating Biological Contactor Process was first used in Germany in the 1920's. In the United States, Doman (2) worked with an experimental plant in Connecticut in 1925. He achieved a 27% BOD_5 removal. Inadequate primary sedimentation and too little surface area per unit reactor volume contributed to the poor results. However, little interest was displayed in the process until the 1960's when Hartmann (3) reported the results of extensive experiments with two RBC plants.

In 1968, Welch (4) used synthetic wastes to evaluate the RBC process. He found that biomass on the disc was equivalent to as much as 17,000 mg/l VSS (Volatile Suspended Solids) when dispersed in mixed liquor.

In 1970, Antonie (5) studied the response of RBC when subjected to fluctuating loading rates. It was found that the RBC performed even better in varying flow pattern than at steady state. He also performed some studies on dairy wastes. The power consumption was only 20-80 hp-hr/1000 lb. BOD_5 removal. The solids settling rate was 15-25 ft/hr and settled sludge contained 4% solids. Birk and Hynek (6) investigated cheese waste treatment and found that the acidic nature of the waste did not cause any problem in the RBC.

In 1971, Torpey (7) attempted to treat wastewater to potable level. He used a 10 stages RBC and removed 93% BOD_5. Six activated

carbon columns were used to further reduce TOC (Total Organic Carbon) from 10 mg/l to 1 mg/l. Disinfection would be required if the effluent was used for domestic or industrial purpose.

Pretorius (8) in an investigation on some operational characteristics of RBC found that bacteria in the first stage were sphaerotilus and beggiatoa where other stages contained mostly nitrate forming bacteria and fungi. He found that COD was removed at a rate of 0.49 g. COD/g. biomass/day and nitrification occured at a rate of 0.138 nitrate-nitrogen/g. biomass/day.

Autotrol Corporation (9) has specified two process design criteria for their RBC unit. They found that a peripheral velocity of 58 ft/min. would give the system the highest removal rate. They also recommended a spacing of 0.5 inch between two discs surfaces.

Chittenden and Wells (10) studied RBC in treating anaerobic lagoon effluent. They anticipated that the first stage of RBC would only convert the waste to aerobic state; however, they were surprised to find that 80% of the BOD_5 was removed in the first stage. They used a higher RPM in the first stage than other stages to increase oxygen transfer.

Cochrane (11) compared RBC with aeration treatment of cannery wastes. It was found that the hydraulic retention time in RBC was only 1 to 5% of the aeration unit for the same BOD_5 removal. The final effluent also contained the same SS (Suspended Solids), and power consumption was much less in RBC. However, sludge from RBC required further treatment.

In 1972, Labella (12) compare the capital and maintenance cost of

RBC, activated sludge and activated lagoon. The costs were base on a 0.4 MGD plant in 1972.

Type of System	Capital Cost	Maintenance Cost (annual)
Activated Lagoon	$240,000	$17,000
Activated Sludge	$175,000	$17,000
RBC	$245,000	$ 9,200

In 1973, Sack (13) ran a RBC to treat wastewater from a summer camp. There was no objectionable odor and maintenace required was only 1.3 hr/week. The removal rates for BOD_5 were 84.5%, COD 71%, TOC 71%, SS 75%, total nitrogen 40% and ammonia 25%.

Antonie (14) in 1973 extended his earlier work using the RBC in different food processing plants. For the same type of RBC, the BOD_5 data he collected were:

Type of Wastewater	Plant Capacity (GPD)	Influent BOD_5 (mg/l)	Effluent BOD (mg/l)
Dairy	200,000	1,000	250
Bakery	50,000	2,000	300
Winery	350,000	700	35
Poultry	130,000	4,500	2,000

In 1974, Gillaspie and William (15) evaluated RBC performance on 11 pilot plants in the lumber industry. With an initial BOD_5 of 500 to 300 mg/l, the BOD_5 removal ranged from 58 to 95%, depending on the kind of waste.

In 1975, Davies and Pretorius (16) used RBC for denitrification.

They found that the optimum removal was obtained at pH=7 and temperatures between 20 to 30^{o}C. Recycling of bacterial was required, and the sludge did not have settling problems due to evolved nitrogen gas.

In 1976, Obayashi (17) studied the usage of RBC in oxidizing ammonia of supernatant from digested sludge lagoons. At 10^{o}C, 99.4% ammonia was removed. At 20^{o}C, 99.8% was removed.

Bintanja (18) studied the differences between using pure oxygen and air in RBC. With pure oxygen, more COD was removed and less sludge with better settling properties was produced. For pure oxygen, substrate was limiting. Torpey (19) also performed a test with oxygen enriched atmospheres. Instead of using pure oxygen, he enriched the first stage of RBC by 60%, and found that BOD_5 removal increased from 34% to 52%.

Finally, Chesner (20) in scale up design of RBC found that using peripheral speed instead of RPM as a means of control would improve the perfomance of RBC.

Mathematical Models for Microbial Growth:

In order to develop the mathematical model for the Rotating Biological Contactor, several mathematical models which represent microbial growth for fixed-film systems were investigated. These models are presented in this section.

In 1950, Monod (21) presented a mathematical expression based on his work with bath reactors. He defined specific growth rate, μ, as the rate of increase of organism concentration per unit concentration

of organisms,

$$\mu = \frac{dx}{dt} / x$$

where x is the organism concentration and μ is the growth rate. Empirically, Monod derived an expression for batch reactor kinetics as follows,

$$\mu = \frac{(\hat{\mu})(C)}{K_c + C}$$

where $\hat{\mu}$ = maximum specific growth rate

C = substrate concentration

K_c = concentration of C at which μ is one-half of $\hat{\mu}$

The use of a saturation function for organism growth kinetics is not universal. Monod did successfully fit his experimental data with this type of function; however, many deviations from it have been noted. Many modifications of the model have been attempted, but it is a reasonable assumption and starting point when modeling dispersed culture systems.

In 1968, Busch and Hughmark (22) found that most fixed-film models were only good for laminar flow. They developed a model by dividing the liquid film into a number of rectangular cells. They used a digital computer to calculate diffusion in each cell. From the experimental data they collected, they discovered that liquid film flow was not laminar. This indicated that eddy diffusion took place in the liquid film.

In 1969, Antonie and Welch (23) modeled RBC in treating dairy

wate. They used dimensional analysis to find a relationship between different systems parameters and system efficiency. By using multiple regression analysis, they derived the following empirical model.

$$X = (K\, B^b C^c D^d)^{\frac{1-(a+1)^N}{1-(a+1)}} (A)^{(a+1)^N - 1}$$

where X = COD removal rate

A = influent COD

N = number of stages in RBC

A,B,C,D = RBC system's parameters

K,a,b,c,d = empirical constants

When the experimental data were compared with the model prediction, it was found that the RBC performed more efficiently than model prediction indicated. They explained that it was due to the fact that the dairy waste contained high colloidal protein and fat which were easily removed by organisms.

In 1970, Sheikh (24) studied the relationship between organic retention time and trickling filter efficiency. Using dimensional analysis, he formulated the following equation:

$$\text{Median time} = K\left(\frac{(Av)(D)}{Q^{0.78}}\right)$$

where Av = specific surface area

D = filter depth

Q = hydraulic loading rate

K = constant

He found that median time or standard deviaions were very useful in

computing filter efficiency.

Monadjemi and Behn (25) in modeling a trickling filter applied mass diffusion theory.

$$\frac{\partial^2 c}{\partial x} = \frac{\partial c}{\partial z}$$

where c = concentration of substrate

x = distance away from air-liquid interface

z = depth in filter

Since velocity of substrate in a liquid film varies (velocity being maximum at the air-liquid interface and zero at the liquid-biofilm interface) so they used an extra term to take care of the velocity gradient.

$$(D)\ \frac{\partial^2 c}{\partial x^2} = (V_{max})\ \left(\ 1 - \frac{x^2}{\delta^2}\ \right)\ \frac{\partial c}{\partial z}$$

where V_{max} = velocity of liquid flowing through the filter

δ = liquid film thickness

D = diffusivity of substrate

From the above equation, they also derived an equation linking efficiency of filter with oxygen uptake rate.

$$\log\ (\ 1 - e\ ) = D \log\ (\ f[k,z]\ )^{\frac{1}{z_o}}$$

where e = efficiency (in fraction)

z_o = filter depth

In 1972, Quirk, Lauler and Matusky (26) developed a model for fixed-film reactors. They only considered the depth and size of the

reactor and the flow rates. They considered the reaction to be first order. The following equations were derived

$$\frac{dL}{dH} = \frac{(K)\ (D)\ (W)}{(Q + R)}$$

$$\frac{L_o}{L_e} = \exp\left[\frac{(K')\ (H)}{(Q + R)}\right]$$

where

L_o = influent BOD_5

L_e = effluent BOD_5

Q = flow rate of untreated influent

R = recirculation rate

K = BOD_5 removal rate

K' = BOD_5 removal rate for first order

D = length

W = width

The model was verified by operating data from municipal sewage, kraft mill, sulfite, hard-board mills and yeast fermentation processes. The model prediction approximated operating data.

Grieves (27) derived the first RBC dynamic model in 1972. Based on this model, he derived a heterogeneous model, a pseudo-homogeneous model and a steady state model. The dynamic model was based on substrates diffusing into biofilm which contained substrate-consuming microorganisms. The substrate consumption was based on Monod's expression. He derived the following non-linear partial differential

equation:

$$\frac{\partial s}{\partial t} = (D)\frac{\partial^2 s}{\partial x^2} - \frac{\hat{\mu} \cdot X}{Y}\left(\frac{s}{K_s + s}\right)$$

where s = substrate concentration

D = diffusivity constant

X = microorganism concentration

$\hat{\mu}$ = maximum specific growth rate

Y = yield coefficient

x = distance inside biofilm

Two more equations were derived to govern the boundary conditions, one for diffusion of substrate in liquid film when disc was exposed in air and another for diffusion of substrate in the bulk liquid when disc was submerged in it. Grieves used digital and analog computers to solve the three equations. The prediction seemed to match his experimental data (using synthetic substrate) in a satisfactory manner; however, the model is only good if substrate is the limiting factor. For high substrate concentration or high flow rate where the system is deprived of oxygen, the model is not applicable.

Kornegay (28) derived two models, one for trickling filters and another for RBC. In both models, equations were derived from mass diffusion for the whole wastewater treatment system. His equation for the trickling filter model is:

$$So - Se = \frac{P}{F}(A)(H)(Z) - Kg \cdot \log_e\left(\frac{So}{Se}\right)$$

where $$P = \frac{1}{Y}\left(\mu_{max}\right) X$$

His equation for the RBC model is:

$$F\left(S_o - S_1\right) = \frac{2 \cdot \mu_{max}}{Y_g}(N)(\pi)(X)(d)\left(r_o^2 - r_u^2\right)\left(\frac{S_1}{K_g + S_1}\right)$$

where S_o = influent concentration

S_e = effluent concentration of trickling filter

S_1 = effluent concentration of RBC

X = microorganism concentration

K_g = Monod's saturation constant

Y_g = growth yield

A = surface area

H = cross-sectional area

Z = depth of filter

F = substrate flow rate

N = number of discs

r_o = disc radius

r_u = disc unsubmerged radius

d = thickness of biofilm

The equation for the RBC was found to be useful only for high and low loading rates. At high loading rate, RBC efficiency was dependent on the product of substrate concentration and flow rate. At low loading rate, efficiency was dependent either on flow rate or substrate concentration. Maximum efficiency occured when discs was 50% submerged.

For the trickling filter, efficiency was higher when filters were placed in series than in parallel. Recycling would increase efficiency

if flow rate was higher than 600 gallon/day/ft^2. For most economical operation, the area to volume ratio should be 27 ft^2/ft^3.

In 1974, Bintanja and Boelhouwer (29) derived an equation for calculating the amount of oxygen transferring from air into liquid film on the RBC. Using mass diffusion equation with known boundary conditions, they solved the equation analytically. The equation is:

$$\frac{\partial C}{\partial t} = (D)\frac{\partial^2 C}{\partial x^2}$$

boundary conditions:

$t = 0$	$0 < x < \delta$	$C = C_o$
$t > 0$	$x = \delta$	$C = C_s$
$t > 0$	$x = 0$	$\frac{\partial C}{\partial x} = 0$

Solving the above equation, they obtained:

$$\frac{C - C_o}{C_s - C_o} = \sum_{n=0}^{\infty} (-1)^n \operatorname{erfc}\left(\frac{(2n+1)\delta - x}{2\,(D \cdot t)^{\frac{1}{2}}}\right) + \sum_{n=0}^{\infty} (-1)^n \operatorname{erfc}\left(\frac{(2n+1)\delta + x}{2\,(D \cdot t)^{\frac{1}{2}}}\right)$$

where C = oxygen concentration in liquid film

C_o = initial oxygen concentration

C_s = oxygen saturation concentration

D = diffusivity of oxygen

t = time

x = distance in liquid film

δ = liquid film thickness

erfc = error function

n = higher number represent higher degree of accuracy in error function

In 1976, McCarty and Williamson (30) derived a model on substrate ultilization in bacterial film. Using Fick's law of diffusion and Monod's expression, they came up with an equation:

$$\frac{\partial^2 S}{\partial z} = \frac{(K)(S)(X)}{D\,(S + K)}$$

They employed Runga-Kutta technique to solve the equation. They concluded that if:

$$S_{oa} = \frac{(D_{cd})(\nu_a)(MW_a)}{(D_{ca})(\nu_d)(MW_d)} \cdot S_{od}$$

then electron acceptor (O_2) would be flux limiting; otherwise the electron donor (substrate) would be flux limiting:

where S = electron concentration

MW = molecular weight

D = diffusivity

ν = viscosity

subscripts: a = electron acceptor

d = electron donor

They also verified their model by their experimental data. The verification showed that prediction was accurate for deep biofilms and when liquid film was stagnant.

Howell and Atkinson (31) developed a model for trickling filters to show the influenced of sloughing on trickling filter BOD_5 removal efficiency. The following parameters were found to affect the filter efficiency.

a) sloughing concentration

b) influent concentration

c) size of filter packing

d) number of filter units

e) time intervals of sloughing

Since sloughing occured discretely rather continuously, they divided the filter into many sections. The equation they derived was based on the rate of growth of film within the jth filter unit is as follows:

$$\rho_o = \frac{dL_j}{dt} = (Y_o)(N)$$

where ρ_o = density of biomass

L_j = substrate concentration

t = time

Y_o = yield coefficient

N = substrate consumption / surface area

The model was used to design filter depth and the optimum loading rate.

Hansford, Andrews and Grieves (31) based on Grieves' dynamic model, derived the following steady-state equation:

$$C_b = \frac{(F)(C_o) + (E_f)(C_1(\beta))}{F + F_f + (K)(A)(S)(\frac{Kl}{Kl + 1})}$$

where C_o = substrate influent concentration

C_b = substrate effluent concentration

C_1 = substrate concentration in liquid film

β = angle of submergence of disc

F = substrate flow rate

F_f = flow rate of liquid film entering RBC

K = mass transfer coefficient in liquid film

A = service area of liquid film

Kl = Monod's saturation constant

Friedman, Robbin and Woods (33) examined the effect of RBC rotational speed on its efficiency. They developed the following equation:

$$K = [(a)(\log_e \theta)(Cin) + (a)(b)] \log_e \omega$$

a = -36.21

b = 228.85

where θ = retention time

Cin = influent concentration

ω = rotational speed

K = substrate removal constant

From the experimental data they collected, disc rotational speed was quite insignificant at low loading rate. However at high loading rate, BOD removal depended significantly on rotational speed. By studying other RBC, they found that most of them had overdesigned and effluent BOD was below effluent standard. This resulted in high wastewater treatment cost.

IV. THEORETICAL CONSIDERATION

Owing to the complexity of the RBC system, the following assumptions were made when developing this mathematical model.

1. There is no change in substrate concentration in the radial or θ direction.
2. The microorganism concentration is assumed to be constant throughout the biofilm.
3. The mass of liquid film adheres to the same biofilm throughout the disc's rotation.
4. There is no substrate removal in the bulk liquid in the reactor.
5. Substrate removal in the liquid film occurs only in diffusion into the biofilm.
6. Diffusivity coefficients of substrate and oxygen are based on water as medum.
7. The model is valid only for steady state condition, since the biofilm thickness is constant.

(1) Mass Balance on Substrate in Biofilm

a) Substrate Accumulation = Flow in by Diffusion - Flow out by Diffusion - Ultilization by Microorganisms

$$\text{Substrate Accumulation} = (\Delta V)\frac{\partial S}{\partial t}$$

$$\text{Flow in by Diffusion} = -\left(D_{23}\cdot A\right)\left.\frac{\partial S}{\partial X}\right|_{x,t}$$

$$\text{Flow out by Diffusion} = -\left(D_{23}\cdot A\right)\left.\frac{\partial S}{\partial x}\right|_{x+\Delta x,t}$$

$$\text{Ultilization by Microorganisms} = \Delta V\left(\frac{\hat{\mu}\cdot X_c}{R}\right)\left(\frac{S}{K_s + S}\right)\left(\frac{U}{K_u + U}\right)$$

$$(\Delta V)\frac{\partial S}{\partial t} = -\left(D_{23}\cdot A\right)\left.\frac{\partial S}{\partial X}\right|_{x,t} + \left(D_{23}\cdot A\right)\left.\frac{\partial S}{\partial x}\right|_{x+\Delta x,t} - \Delta V\left(\frac{\hat{\mu}\cdot X_c}{R}\right)\left(\frac{S}{K_s + S}\right)\left(\frac{U}{K_u + U}\right)$$

where

S = concentration of substrate

U = concentration of oxygen

x = distance into biofilm from liquid film and biofilm interface

ΔV = volume of biofilm submerged in bulk liquid (in reactor)

A = submerged area in the plane perpendicular to the direction of diffusion

t = time

D_{23} = diffusivity of substrate

Ω = maximum specific growth rate

X_c = microorganism density

R = yield coefficient = $\frac{\text{mass of organisms produced}}{\text{mass of substrate consumed}}$

K_s = Monod's saturation concentration of substrate

K_u = Monod's saturation concentration of oxygen

Since $\Delta V = A \cdot (\Delta x)$

hence
$$(A \cdot \Delta x) \frac{\partial S}{\partial t} = (A \cdot D_{23}) \left[\frac{\partial S}{\partial x} + \frac{\partial^2 S}{\partial x^2}(\Delta x) - \frac{\partial S}{\partial x} \right]$$
$$-(A \cdot \Delta x) \left(\frac{\Omega \cdot X_c}{R} \right) \left(\frac{S}{K_s + S} \right) \left(\frac{U}{K_u + U} \right)$$

and
$$\frac{\partial S}{\partial t} = (D_{23}) \frac{\partial^2 S}{\partial x^2} - \left(\frac{\Omega \cdot X_c}{R} \right) \left(\frac{S}{K_s + S} \right) \left(\frac{U}{K_u + U} \right)$$

b) For Oxygen Ultilization,

$$\frac{\partial U}{\partial t} = (D_{13}) \frac{\partial^2 U}{\partial x^2} - (Y) \left(\frac{\Omega \cdot X_c}{R} \right) \left(\frac{S}{K_s + S} \right) \left(\frac{U}{K_u + U} \right)$$

where D_{13} = diffusivity of oxygen

Y = stoichiometric coefficient

$\left[= \frac{\text{mass of oxygen consumed}}{\text{mass of substrate produced}} \right]$

With Initial Condition,

$S = 0$ (except at boundary)

$$U = 0$$

$$S_b = S_o$$

$$S_f = S_o$$

and Boundary Condition,

$$\left.\frac{\partial U}{\partial x}\right|_{x=L} = 0$$

$$\left.\frac{\partial S}{\partial x}\right|_{x=L} = 0$$

at $x = 0$ $\left.\begin{array}{l} U = U_b \\ S = S_b \end{array}\right\}$ in bulk liquid

at $x = 0$ $\left.\begin{array}{l} U = U_f \\ S = S_f \end{array}\right\}$ when exposed to air

where S_o = concentration of substrate in influent

S_b = concentration of substrate in bulk liquid

S_f = concentration of substrate in liquid film

U_b = concentration of oxygen in bulk liquid

U_f = concentration of oxygen in liquid film

L = thickness of biofilm

(2) Mass Balance in Liquid Film

$$(\Delta V_f)\frac{\partial S_f}{\partial t} = (K_L \cdot A')\frac{\partial S}{\partial x}$$

$$(\delta \cdot A') \frac{\partial S_f}{\partial t} = - (K_L \cdot A') \left. \frac{\partial S}{\partial x} \right|_{x=0}$$

$$\frac{\partial S_f}{\partial t} = - \left(\frac{K_L}{\delta} \right) \left. \frac{\partial S}{\partial x} \right|_{x=0}$$

where δ = thickness of liquid film

K_L = substrate transfer coefficient from liquid film to biofilm

A' = area in the plane perpendicular to the direction of duffusion when disc is exposed to the air

V_f = volume of liquid film

With Initial Condition,

$$S_f = S_b \qquad \text{at } \theta = 0$$

$$\frac{\partial S_f}{\partial x} = 0$$

where θ = angular direction

(3) Mass Balance of Oxygen in Liquid Film

$$\frac{\partial U_f}{\partial t} = (D_{12}) \frac{\partial^2 U_f}{\partial x^2}$$

With Boundary Condition,

$$t = 0 \qquad 0 < x < \delta \qquad U = U_o$$

$$t > 0 \qquad x = \delta \qquad U = U_s$$

$$t > 0 \qquad x = 0 \qquad \frac{\partial U}{\partial x} = 0$$

The following equation is obtained when solved analytically (29).

$$\frac{U_f - U_o}{U_s - U_o} = \sum_{n=0}^{\infty} (-1)^n \cdot \text{erfc} \left[\frac{(2n+1)\delta - x}{(2)(D_{12} \cdot t)^{\frac{1}{2}}} \right]$$

$$+ \sum_{n=0}^{\infty} (-1)^n \text{erfc} \left[\frac{(2n+1)\delta - x}{(2)(D_{12} \cdot t)^{\frac{1}{2}}} \right]$$

Since only concentration at x = 0 is required, hence

$$\frac{U_f - U_o}{U_s - U_o} = (2) \sum_{n=0}^{\infty} (-1)^n \text{erfc} \left[\frac{(2n+1)(\delta)}{(2)(D_{12} \cdot t)^{\frac{1}{2}}} \right]$$

$$U_f = U_o + (2)(U_s - U_o) \cdot$$

$$\sum_{n=0}^{\infty} (-1)^n \text{erfc} \left[\frac{(2n+1)(\delta)}{(2)(D_{12} \cdot t)^{\frac{1}{2}}} \right]$$

where $U_s = \dfrac{\text{Partial Pressure}}{\text{Henry's constant}_{\text{oxygen}}}$

U_o = initial oxygen concentration of liquid film

(4) Mass Balance in the Completed Mixed Reactor

Substrate Accumulated in Bulk Liquid per unit time = Substrate Mass Flow Rate of Influent - Substrate Mass Flow Rate of Effluent - Loss of Substrate to Liquid Film per unit time - Loss of Substrate to Biofilm (diffusion) per unit time

$$\text{Substrate Accumulated in Bulk Liquid per unit time} = (V_b)\frac{\partial S_b}{\partial t}$$

$$\text{Substrate Mass Flow Rate of Influent} = (F)\,S_o$$

$$\text{Substrate Mass Flow Rate of Effluent} = (F)\,S_b$$

$$\text{Loss of Substrate to Liquid Film per unit time} = (F_f)\,S_f(\theta=0) - (F_f)\,S_f(\theta=\beta)$$

$$\text{Loss of Substrate to Biofilm per unit time} = (K_L)(A)\left.\frac{\partial S}{\partial x}\right|_{x=0}$$

where V_b = volume of bulk liquid

F = substrate flowrate

F_f = liquid film flowrate

hence
$$(V_b)\frac{\partial S_b}{\partial t} = (F)(S_o) - (F)(S_b) - (F_f)\,S_f(\theta=0) + (F_f)\,S_f(\theta=\beta) - (K_L)(A)\left.\frac{\partial S}{\partial x}\right|_{x=0}$$

With Initial Condition:

$$S_b = S_o$$

$$S_f(\theta=0) = S_b$$

$$(V_b)\frac{\partial S_b}{\partial t} = (F)(S_o) - (F)(S_b) - (F_f)(S_b) + (F_f)(S_f) - (K_L)(A)\left.\frac{\partial S}{\partial x}\right|_{x=0}$$

In the mathematical model, there are two non-linear parabolic partial differential equations (Eq. 1a,b) which are required to be solved simultaneously. In solving partial differential equations, there are generally 3 methods. They are the forward difference method, the backward difference method and the Crank-Nicolson method.

The first two methods are only first-order correct, and the solution will not be very precise. The forward difference method also has a disadvantages in that the ratio $\Delta t/(\Delta x)^2$ must remain less than or equal to $\frac{1}{2}$. This restriction is a rather serious one, for in order to minimize the truncation error, the size of Δx must be small. This means that Δt has to be small too, and it would require a much

longer computer time to obtain results with the same degree of accuracy.

To obtain a second-order-correct analog for du/dt, the Crank-Nicolson method (34) was used. In order to use Crank-Nicolson method, the boundary conditions must be known. To find the relationship among boundary points, three more equations (Eq. 2,3,4) were derived for the model and the boundary conditions for each new time level were calculated from them. Since they were ordinary differential equations, a simplier method was employed to solve them. In order to stay as second-order-correct, the Improved Euler Method was used. A computer program in Fortran language was written to solve these equations. It is shown in Appendix A.

Parameter Value Selection:

Before the mathematical model can be solved, values for the parameters to be used in the model must be known. The parameters of importance are as follows:

L, δ - thickness of biofilm and liquid film, respectively. The thickness of biofilm was measured by inserting a piece of thin glass into the biofilm after it had been dried for 30 minutes to make sure that the liquid film was gone. The glass was then placed under a microscope which was scaled in microns, and the thickness of the film was measured. Ten measurements were made and the average film thickness was found to be 150 microns. For liquid film thickness, Hartmann (27) is the only investigator to have measured the average thickness of the liquid film. He accompolished this by carefully positioning a

scraper so that it just touched the organism film surface on the rotating disc. Liquid film was collected continuously for a known period of time, and Hartmann found the average thickness to be about 40 microns. This technique was tried but would not be reproduced, with results ranging from 50 to 200 microns. In this case, we chose an average value of 150 microns in this simulation.

μ - maximum specific growth rate. From pertinent literature, μ was found to have a value from 0.2/hr. to 0.54/hr. for a dispersed culture system growing on glucose. Korenegay (35) obtained a value of 0.28/hr. for a fixed film system with glucose as a substrate, thus this value was used.

Y - yield coefficient. Y is usually taken as a constant for a particular organism-substrate system. However, Y may not be a constant for transient conditions. Both Blackwell (36) and Young (37) have demonstrated that the value for Y can approach unity during periods of transient operation. At steady state Y is of the order of 0.26 (37) to 0.64 (36). An average value, 0.40, was taken for this simulation.

X - density of active mass in the biofilm. Hoehn (38) showed that X varied from about 20 to 105 mg/ml. In this case, 20 mg/ml was chosen because the biofilm did not appear to be very dense.

K1, K2 - Monod Saturation Coefficient, for oxygen and substrate, respectively. For a single organism in a dispersed culture, K2 is of the order of 4 to 10 mg/l glucose (39), which is the value that Monod originally reported. Powell (40) has shown mathmatically that for an organism in the dispersed state, the value of K2 can be inflated by

the effect of diffusion of substrate across the cell membrane. It can be visualized that the apparent value of K2 for an agglomerate of organisms, a floc, will be appreciably higher than the value of a dispersed organism because of the inclusion of the effect of diffusion. Kornegay (35) reported a value of 80 mg/l glucose for his fixed-film reactor. In this case, a middle value of 80 mg/l was chosen for simulation. K1 has been found to be approximately equal to 1 mg/l.

D_{13}, D_{23} - diffusivity of oxygen and substrate respectively. The diffusivity of oxygen had been measured and reported values differ by as much as an order of magnitude. Tomlinson and Shaddon (41) showed that the diffusivity of oxygen varied from 1.5×10^{-5} cm^2/sec to 22.0×10^{-5} cm^2/sec depending upon the physiological state of the film and the nutrient limiting growth. The latter value was for a loosely packed, predominantly fungal film. This is to be compared with a value of 2.5×10^{-5} cm^2/sec for oxygen diffusing in pure water. Thus for oxygen, a value of 5.0×10^{-5} cm^2/sec was chosen for simulation (42). For substrate, the diffusivity was chosen to be the value of glucose diffusing in pure water, 0.64×10^{-5} cm^2/sec at $20^{o}C$ (42) since pertinent literature could not be found for diffusivity of substrate in fixed film.

K_L - liquid film transfer coefficient across the liquid-solid interface. No data can be found for the transfer coefficient in biological system such as one under consideration. However, Danckwerts (43) presents curves for variation of the liquid film coefficient for different media used in packed towers in the chemical engineering industry. For the absorption of CO_2 in water, values varied from 0.4

to 2.2 X 10^{-2} cm/sec when flow rate was varied. In this case, a middle value, 1.0 X 10^{-2} cm/sec was used in simulation.

He, P - Henry's constant and partial pressure for oxygen respectively. Henry's constant for oxygen is 37000 atm./mole fraction at 60^{o}F and 45500 atm./mole fraction at 80^{o}F (44). The average room temperature in the laboratory was approximately 78^{o}F in summer, so 45,500 was chosen as the value in simulation. The partial pressure for oxygen is 0.21 atm. (44).

V. EXPERIMENTAL RESULTS

Data were collected from a pilot plant over a period of three months. The following parameters were measured.

1. Five Days Biological Oxygen Deman, BOD_5 (shown in Fig. 2)
2. Chemical Oxygen Demand, COD (shown in Fig. 3)
3. Ammonia-nitrogen, NH_3-N (shown in Fig. 4)
4. Nitrate-nitrogen, NO_3-N (shown in Fig. 5)
5. Nitrite-nitrogen, NO_2-N
6. Phosphate, PO_4
7. Total Suspended Solids, TSS (shown in Fig. 6)
8. Dissolved Oxygen in RBC bulk liquid, D.O.
9. pH (shown in Fig. 7)

For phosphate, nitrite and dissolved oxygen, data are shown in Appendix B.

Influent samples were taken from the barrel containing sewage collected from the Westwood Boulevard sewer at the UCLA campus. The barrel was used as a primary settling tank. In order to make sure that all the large solids had settled to the bottom of the barrel, samples were not taken until two hours after collection. Samples for various stages were collected from the biodisc. In order to study the effect of clarification of RBC effluent, a 8" diameter funnel was used as a clarifier.

Before the beginning of data collection, the RBC was allowed to run for three weeks to make sure that microorganisms growth on the

discs was well established and the system had reached a steady state. After those three weeks, the first stage was fully covered by a layer of brown biofilm and the effluent was clear. In the beginning of the fourth week, collection of data was ready to begin; however, the biofilm began to slough off. The color of biofilm became lighter and lighter, and after a few days, only a thin biofilm was left on the discs. After contacting the University Recreation Department, it was found that the university swimming pool was drained because many swimmers suffered from red-eye illness, an indication of water contamination in the pool. Since water in the pool contained quite a large amount of residual chlorine, it was suspected that the chlorine mixed with the sewage could have upset the RBC. It was decided that the experiment had to start all over, so collection of data was postponed for two weeks until a thicker biofilm had grown on the discs. On the sixth week, collection of data began as the disc was covered by a thicker biofilm.

BOD_5 was measured twice a week. The same sample was also analyzed for COD and TSS which will be discussed later. BOD_5 data is presented in Figure 2. Influent BOD_5 had an average value of 150 mg/l which would be considered to be a weak domestic stage (45). BOD_5 in the first stage had an average value of 30 mg/l, indicating a reduction rate of 80%. BOD_5 in the second stage, third stage, effluent and clarified effluent had an average value of 15 mg/l, 10 mg/l, 7 mg/l and 3 mg/l respectively. Since the loading rate was low, so the system was substrate limiting and a fully developed biofilm could not be formed in the latter stages. The loading rate was 1 gallon/hour/ft^2,

providing a retention time of 37 hours, since the RBC had a total volume of 37 gallons.

COD tests were run with the same samples as BOD_5. Usually samples were stored in a refrigerator for 1 day because there was not enough time to run BOD_5 and COD test on the same day. The COD data of the influent, effluent, clarified effluent and soluble effluent were collected. COD data were used to compare with BOD_5 data. If there was a relationship between the two parameters, it could be possible to just collect COD data and estimate BOD_5 data from them. It would save more time since BOD_5 test is longer than COD test. In this case, COD/BOD_5 ratio was found to vary between 1.5 to 2.5.

TSS data were collected twice a week. It was found that the RBC reduced the TSS of influent from an average concentration of 73 mg/l to 32 mg/l, a reduction of 56%. This reduction of solids is advantageous since sludge handling is expensive. Data were also collected for clarified effluent after it had been settled for one hour. It was found that over 98% of the solids were settled, leaving only 1% to 2% of solids with an average concentration of 5.4 mg/l suspending in the effluent.

Ammonia-nitrogen and nitrate-nitrogen data were collected five days a week. Data showed that the sewage contained large amounts of organic nitrogen. Ammonia-nitrogen of sewage increased by more than 100% after sewage had stayed in the storage barrel for 23 hours. During that period of time, organic nitrogen was converted to ammonia-nitrogen anaerobically due to a lack of oxygen in the barrel. In the RBC, nitrification was essentially completed after the first stage.

The concentration of nitrate-nitrogen remained almost the same throughout the rest of the stages. When ammonia was nitrified to nitrate, hydrogen ions were given off to the bulk liquid, thus lowering the pH of the liquid. In this experiment the sewage had a high ammonia concentration, therefore pH control was required to prevent the pH of effluent from dropping below a value of 6 (Minimum pH for effluent discharge). Sodium carbonate was used as the pH buffer. The amount of sodium carbonate added varied daily, depending upon the ammonia concentration of the sewage.

Analysis for nitrite-nitrogen was performed only twice. It was found that only a very small amount of nitrite-nitrogen was present in the first stage of the RBC. For the rest of the stages, nitrite-nitrogen was almost non-existent. Phosphate was measured only once. It was found that there was enough phosphate in the sewage (5.68 mg/l) to substain microorganism growth.

Dissolved oxygen test was performed twice in order to find out how much oxygen was present in different stages. These values were used in the mathematical model computation. The average oxygen concentration for the first stage was 6 mg/l. For the rest of the stages, values ranged from 8 mg/l to 9 mg/l.

Ambient temperature ranged from 22-27^{o}C while temperature in the RBC ranged from 16-19^{o}C. This reduction of temperature was due to the evaporation of water from the RBC.

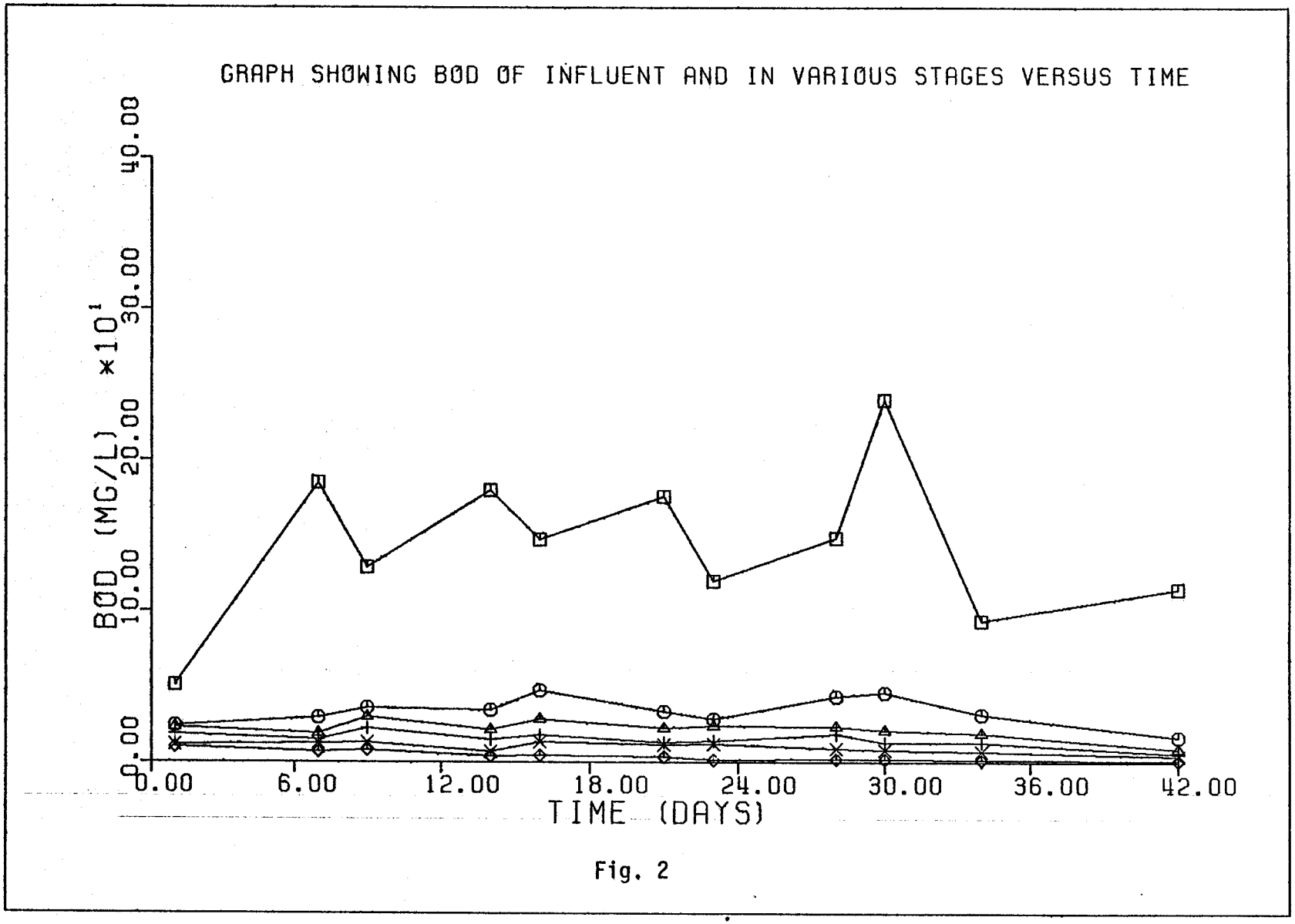
GRAPH SHOWING BOD OF INFLUENT AND IN VARIOUS STAGES VERSUS TIME
BOD (MG/L) *10¹
0.00
10.00
20.00
30.00
40.00
0.00
6.00
12.00
18.00
24.00
30.00
36.00
42.00
TIME (DAYS)

Fig. 2

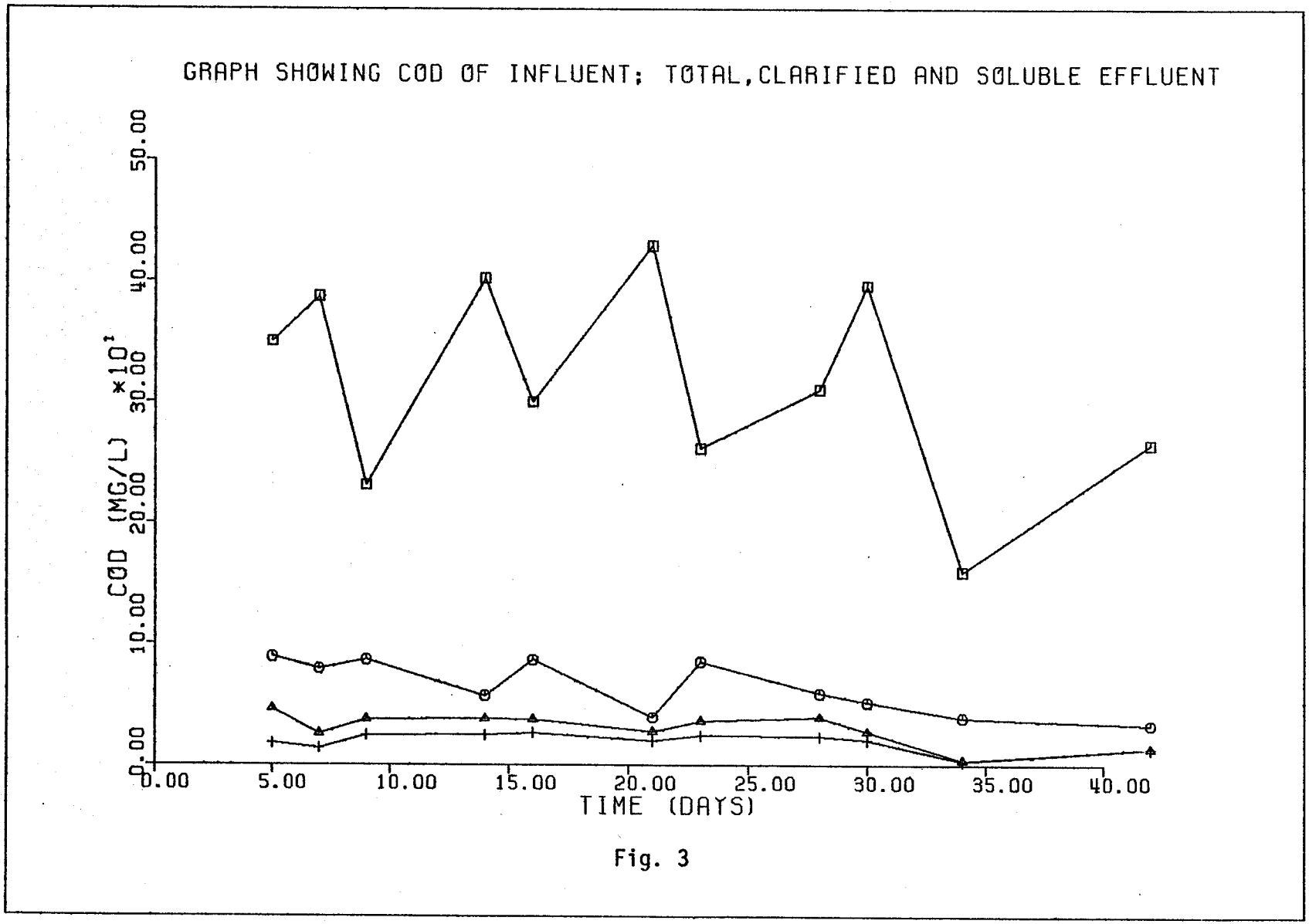
GRAPH SHOWING COD OF INFLUENT; TOTAL,CLARIFIED AND SOLUBLE EFFLUENT
COD (MG/L) *10¹
0.00
10.00
20.00
30.00
40.00
50.00
0.00
5.00
10.00
15.00
20.00
25.00
30.00
35.00
40.00
TIME (DAYS)

Fig. 3

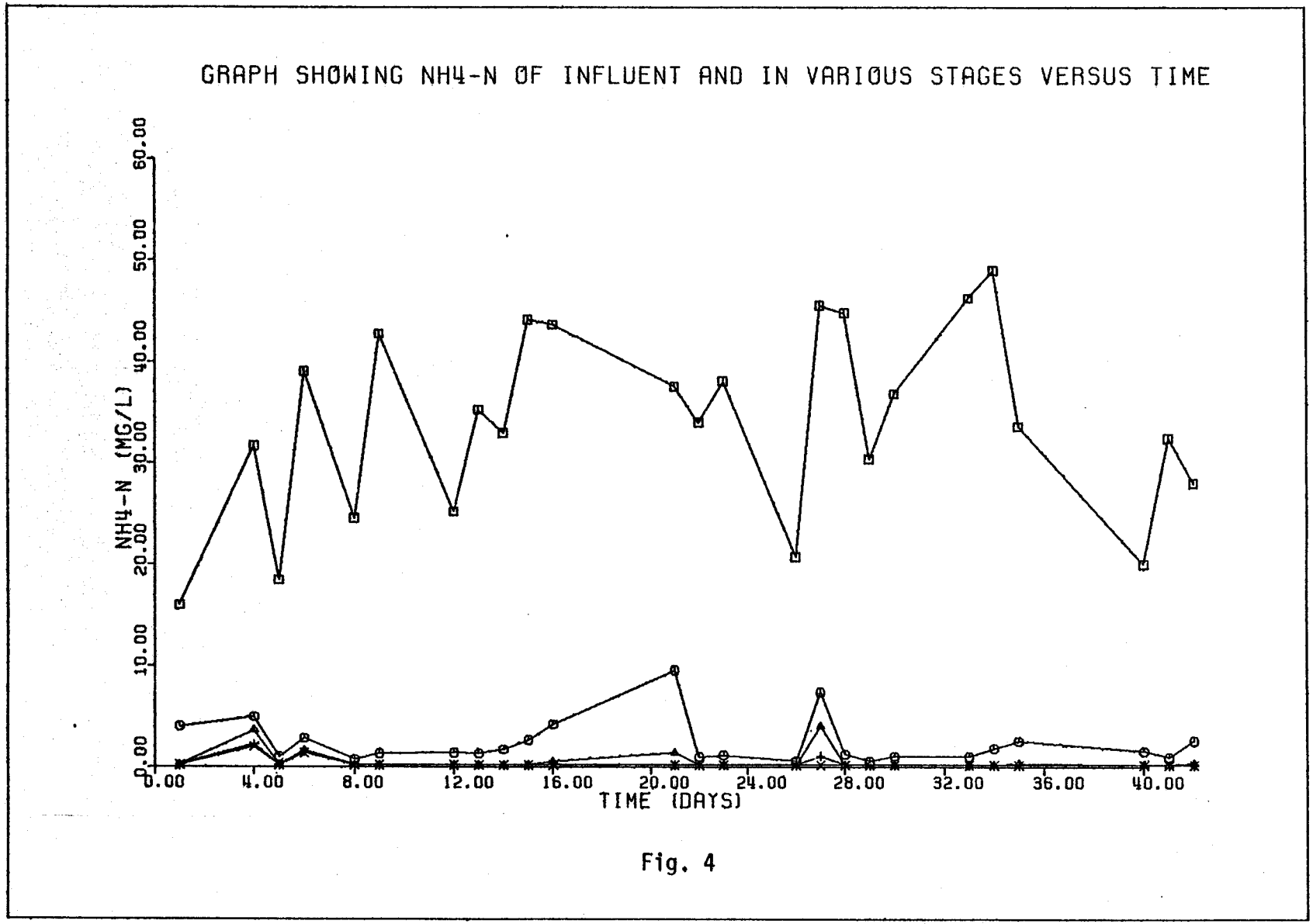
GRAPH SHOWING NH4-N OF INFLUENT AND IN VARIOUS STAGES VERSUS TIME
NH4-N (MG/L)
0.00
10.00
20.00
30.00
40.00
50.00
60.00
0.00
4.00
8.00
12.00
16.00
20.00
24.00
28.00
32.00
36.00
40.00
TIME (DAYS)

Fig. 4

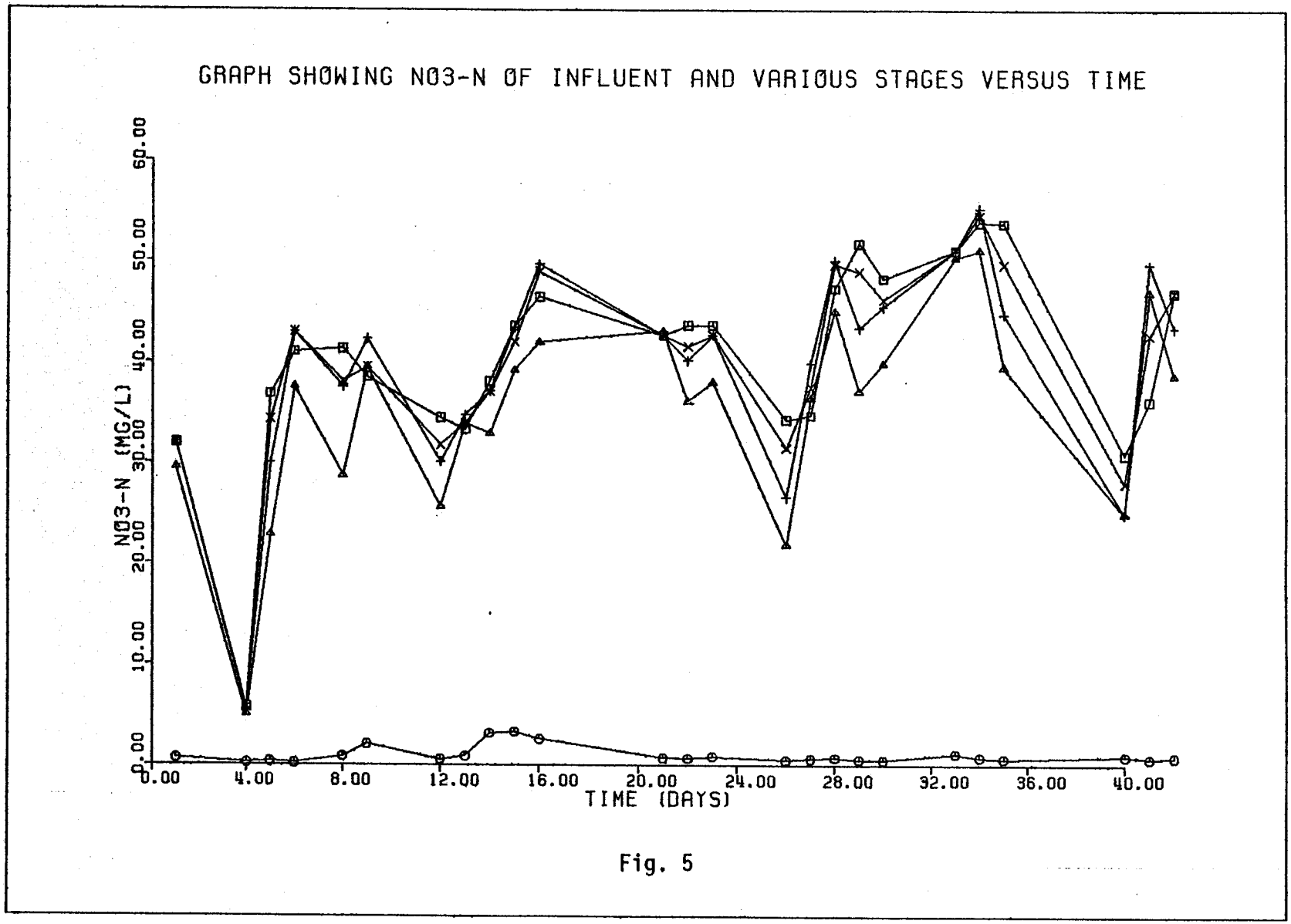
GRAPH SHOWING NO3-N OF INFLUENT AND VARIOUS STAGES VERSUS TIME
NO3-N (MG/L)
0.00
10.00
20.00
30.00
40.00
50.00
60.00
0.00
4.00
8.00
12.00
16.00
20.00
24.00
28.00
32.00
36.00
40.00
TIME (DAYS)

Fig. 5

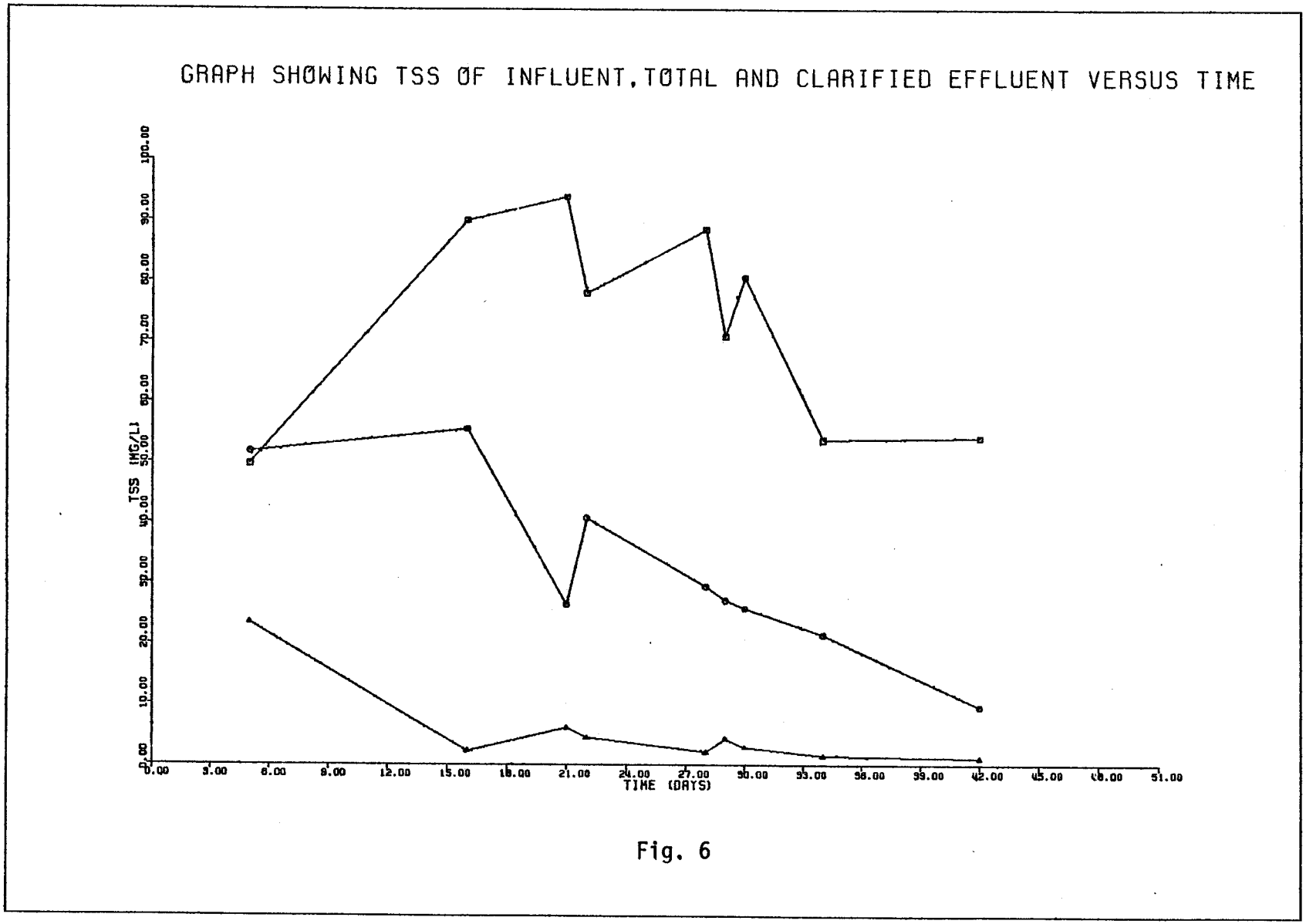
GRAPH SHOWING TSS OF INFLUENT, TOTAL AND CLARIFIED EFFLUENT VERSUS TIME
TSS (MG/L)
TIME (DAYS)

Fig. 6

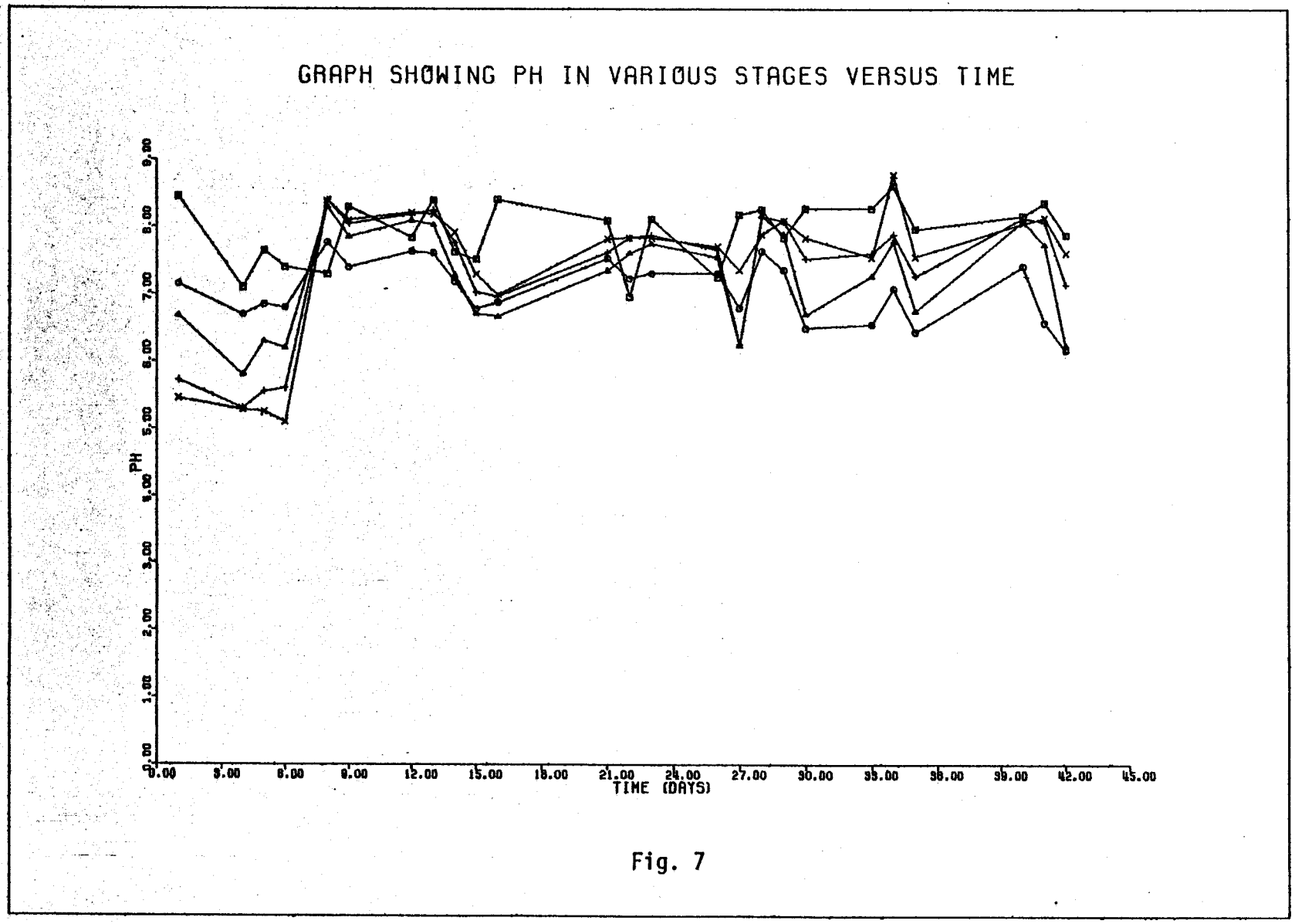
GRAPH SHOWING PH IN VARIOUS STAGES VERSUS TIME
PH
TIME (DAYS)
0.00
3.00
6.00
9.00
12.00
15.00
18.00
21.00
24.00
27.00
30.00
33.00
36.00
39.00
42.00
45.00

Fig. 7

VI. ENGINEERING SIGNIFICANCE

The RBC is an efficient method of treating wastewater because the system is simple to maintain and does not consume as much energy as other methods. Although the capital investment may be expensive, however, the cost can be offset by savings in maintenance and operation. The RBC is particularly favorable for small communities where there is a shortage of skillful operators.

To demonstrate how a small community can benefit economically by treating their wastewater with the RBC, consider the following example. A small town with 4,000 people has a sewage generation rate of 100,000 gallon/day and the BOD_5 of sewage in 250 mg/l. The town is required to reduce their BOD_5 by 92%.

The first step is to calculate the cost of operating an activated sludge plant in this town. The following parameters representing a municipal activated sludge plant are used for calculation.

Mean cell residence time	θ_c	= 10 days
Growth yield	Y	= 0.65 lb. cells/ lb. BOD_5 ultilized
Microorganism decay coefficient	K_d	= 0.1/day
Concentration of microorganisms	X	= 3500 mg/l

The reactor volume (V) is found by:

$$(X)\ (V) = \frac{(Y)\ (Q)\ (\theta_c)\ (\ S_{in} - S_{out}\ )}{1 + (K_d)\ (\theta_c)}$$

$$(3500)\ (V) = \frac{(0.65)\ (100{,}000)\ (10)\ (250\text{-}20)}{1 + (0.1)\ (10)}$$

$$V = 21{,}357 \text{ gallons}$$

where,

Q = Flow rate of sewage (influent)

S_{in} = BOD_5 of influent

S_{out} = BOD_5 of effluent

$\frac{dx}{dt}$ = sludge production rate

$$\frac{dx}{dt} = \frac{(X)\ (V)}{\theta_c} = \frac{(3{,}500)\ (21{,}357)}{10}\ (8.34 \times 10^{-6})$$

$$= 62.3 \text{ lb / day}$$

$\frac{dO_2}{dt}$ = oxygen requirement

$$\text{Oxygen Requirement Per day} = (\text{Food } (BOD_5) \text{ Ultilized Per day}) - (1.42)(\text{Organisms Wasted Per Day})$$

$$+ (4.5)\,(\text{Oxygen for Nitrification Per day})$$

$$\frac{dO_2}{dt} = \frac{dF}{dt} - (1.42)\frac{dx}{dt} + (4.5)\frac{dN}{dt}$$

$$= \frac{(250 - 20)(100{,}000)}{0.68}(8.34\times10^{-6}) - (1.42)(62.3)$$

$$+ (4.5)(100{,}000)(25)(8.34\times10^{-6})$$

$$= 288 \text{ lb } O_2/\text{day}$$

From literature (47), power requirement for commercial coarse bubble diffuser (less maintenance than fine bubble) is 0.5 lb O_2/hp-hr.

$$\text{Power required} = (288) \text{ lb } O_2/\text{day} \times \left(\frac{1}{24}\right) \text{day/hr.} \times \left(\frac{1}{0.5}\right) \text{hp-hr/lb } O_2$$

$$= 24 \text{ hp}$$

Also the recycle pump and the pump from the aeration tank to the clarifier would each require a ½ hp motor. The total power required therefore is 25 hp.

The second step is to calculate the energy requirement of RBC. From the Biosurf Design Handbook from the Autotrol Corp. (46), to obtain a BOD_5 removal rate of 92% with an initial sewage concentration

of 250 mg/l, the hydraulic loading rate should be 3 GPD/FT^2. The power required for 3 GPD/FT^2 is 20 hp/MGD. In this case study, the treatment capacity is only 0.1 MGD, so the power required is 2 hp.

The RBC only requires 2 hp to operate while the activated sludge plant requires 25 hp, therefore there is a rather substantial saving in energy. Furthermore, the RBC requires less labor for maintenance. The following table is a summary of operation and estimated maintenance costs.

Type of Activity	Activated Sludge Plant	Manpower per week	Biodisc	Manpower per week
Sludge Handling	1. Transportation to landfill site	7	None	
	2. Spread on land	1		
Tests	Effluent Analysis	3	Same	3
Control Tests		9	None	
Maintenance	1. Compressor	1	1. Shaft motor	1
	2. Recycle pump	1		
	3. Clarifier	2		
Electric Power	Pump	25 hp	Motor	2 hp

Annual Savings = Operation and Maintenance Cost of ASP per year

- Operation and Maintenance Cost of Biodisc per year

= (24 - 4) manhour X \$5/manhour X 53 weeks/year

+ (25 - 2) hp X 365 days/year X 24 hr/day X \$.023/hphr

= \$5300 + \$4610

= \$9910 / year

Since bio-surf has a higher surface to volume ratio, therefore it is more efficient in treating wastewater. With a loading rate of 3 GPD/FT^2, an annual savings in energy is \$9910/year. In this experiment, a low loading rate was used because the RBC pilot plant has a low surface to volume ratio. Furthermore, the goal of this experiment was to achieve a higher reduction of BOD_5 and a substantial reduction of sludge production. Other experiments have to be done in order to evaluate the economical operation of RBC.

VII. CONCLUSION AND DISCUSSION

1. The RBC is an efficient method of treating wastewater because of its simplicity to maintain and operate, low energy consumption, ability to withstand shock or toxic load, freedom from odors and good sludge settling properties.

2. RBC energy consumption is equivalent to or less than extended aeration activiated sludge plants, and it requires less maintenance and operational skill.

3. For small wastewater treatment plant, the capital cost of RBC is lower than activiated sludge plant; therefore, RBC can result in more savings for small communities.

4. The mathematical model developed in this report has proved to be quite successful. Owing to the low loading rate of sewage, biofilm was fully developed only in the first stage. Therefore, only data in this stage was used to verify the model. The influent with an initial BOD_5 of 250 mg/l was predicted by the model to be reduced to a BOD_5 of 40 mg/l in the first stage, a reduction of 84%. The average BOD_5 in the first stage collected over a period of 42 days was 33.7 mg/l, a reduction of 86.5%. The difference between the prediction and the experimental data is only 2.5%. Since there is insufficient data for other loading rates, so it is not possible to verify the model for other loading rates. As a first step toward quantative analysis of RBC, this model is

considered to be very succrssful. However, more work should be done to include;

(a) diffusion in the radial direction since peripheral velocities for sections having different radial distances from the center of the disc are not the same.

(b) diffusion in the bulk liquid of the RBC since the concentration of oxygen and substrate are not constant in the bulk liquid.

(c) diffusion in the liquid film since there is a concentration gradient across the film (in this model, the gradient is considered to be zero).

(d) gradual changes of the biofilm thickness during its development since fluctuating load can cause the thickness of biofilm to change.

5. Careful selection of parameters for model is necessary.

6. Scale-up of the mathematical model should be performed in order that the model can be applied to other RBC wastewater treatment processes with greater treatment capacities.

APPENDIX A

COMPUTER PROGRAM

FOR THE RBC MATHEMATICAL MODEL

```
C.. REX T. CHAN
C.. UCLA  WATER QUALITY CONTROL LABORATORY, MAY,1978
C.. SUBROUTINE DDUMP DOES THE PRINTING AND PLOTTING
C.. SUBROUTINE DPLOT IS THE CALCOMP PLOTTER
C.. SUBROUTINE TA SOLVES THE SIMULTANEOUS EQUATIONS BY THE THOMAS
C.. ALGORITHM
C.. FUNCTION BOUND1 SETS THE SCALED ENTRANCE BOUNDARY VALUES FOR U AND
C.. S WHEN ALPHA<ANGLE<=360 (DEGREES)
C.. FUNCTION BOUND2 SETS THE SCALED ENTRANCE BOUNDARY VALUES FOR U AND
C.. S WHEN ALPHA>=ANGLE>0 (DEGREE)
C.. FUNCTION SB SETS THE SUBSTRATE BOUNDARY CONDITION OF BIOFILM IN THE
C.. BULK LIQUID
C.. FUNCTION SL SETS THE SUBSTRATE BOUNDARY CONDITION OF BIOFILM IN THE
C.. LIQUID FILM
C.. FUNCTION UL SETS THE OXYGEN BOUNDARY CONDITION OF BIOFILM IN THE
C.. LIQUID FILM
C.. FUNCTION ERFC IS THE ERROR FUNCTION FOR CALCULATING OXYGEN TRANSFER
C.. THE A,B,C,D,AS,BS,CS,DS, ARRAYS ARE THE COEFFICIENT ARRAYS FOR THE
C.. THOMAS ALGORITHM
C.. C1,C2,------ ARE CONSTANTS ASSOCIATED WITH THE PROGRAM
C.. X      =DISTANCE AWAY FROM BOUNDARY (STAGNANT FILM)              (CM)
C.. IR     =NUMBER OF GRID POINTS
C.. DX     =DISTANCE INCREMENT                                       (CM)
C.. DT     =TIME INCREMENT                                          (SEC)
C.. DDT    =TIME INCREMENT FOR THE LAST TIME STEP BEFORE
C..          REENTERING THE BULK FLUID                              (SEC)
C.. FINTIM=FINISH TIME                                              (SEC)
C.. S      =CONCENTRATION OF SUBSTRATE IN BIOFILM                (MG/CM**3)
C.. U      =CONCENTRATION OF OXYGEN IN BIOFILM                   (MG/CM**3)
C.. TS,TU  =TRIAL VALUES FOR ITERATIONS                          (MG/CM**3)
C.. G      =ARRAY CONTAINING THE PROJECTED VALUES OF S           (MG/CM**3)
C.. W      =ARRAY CONTAINING THE PROJECTED VALUES OF U           (MG/CM**3)
C.. SKL    =SUBSTRATE TRANSFER COEFFICIENT BETWEEN LIQUID
C..         FILM AND BIOFILM                                       (CM/SEC)
C.. DIFF1  =DIFFUSIVITY OF OXYGEN                                (CM**2/SEC)
C.. DIFF2  =DIFFUSIVITY OF SUBSTRATE                             (CM**2/SEC)
C.. BMU    =MAXIMUM MICROORGANISM REACTION RATE                     (1/SEC)
C.. BMICRO=MICROORGANISM CONCENTRATION                           (MG/CM**3)
C.. STOIC  =STOICHIOMETRIC COEFFICIENT OF OXYGEN            (MG O2/MG BOD5)
C.. RMS    =MICROORGANISM/FOOD(SUBSTRATE) RATIO
```

```
C.. B1    =OXYGEN CONCENTRATION AT WHICH OXYGEN REACTION
C..        RATE IS ONE-HALF                                    (MG/CM**3)
C.. B2    =SUBSTRATE CONCENTRATION AT WHICH MU IS ONE-HALF     (MG/CM**3)
C.. THICK =BIOFILM THICKNESS                                      (CM)
C.. TLF   =THICKNESS OF LIQUID FILM                               (CM)
C.. R     =RADIUS OF DISC                                         (CM)
C.. DEPTH =DEPTH OF SUBMERGING DISC                               (CM)
C.. AREA  =SURFACE AREA OF DISC IN ONE STAGE                    (CM**2)
C.. PER   =PERCENT OF DISC SUBMERGED IN BULK LIQUID             (%/100)
C.. ALPHA =ANGLE RELATING TO THE DEPTH OF DISC SUBMERGENCE      (RADIAN)
C.. THETA =ANGLE OF ROTATION IN THE UNWETTED AREA               (RADIAN)
C.. OMEGA =ROTATIONAL VELOCITY                                    (RPM)
C.. P     =OXYGEN PARTIAL PRESSURE                           (ATMOSPHERE)
C.. HE    =HENRY'S CONSTANT                          (ATM/MOLE FRACTION)
C.. SB    =CONCENTRATION OF SUBSTRATE IN BULK LIQUID            (MG/CM**3)
C.. SL    =CONCENTRATION OF SUBSTRATE IN BIOFILM WHEN
C..        EXPOSED TO AIR                                      (MG/CM**3)
C.. SO    =CONCENTRATION OF SUBSTRATE IN INFLUENT              (MG/CM**3)
C.. UB    =CONCENTRATION OF OXYGEN IN BULK LIQUID              (MG/CM**3)
C.. UL    =CONCENTRATION OF OXYGEN IN LIQUID FILM WHEN
C..        EXPOSED TO AIR                                      (MG/CM**3)
C.. US    =OXYGEN SATURATION CONCENTRATION                     (MG/CM**3)
C.. SMAX  =MAXIMUM CONCENTRATION OF S USED FOR PLOTTING        (MG/CM**3)
C.. UMAX  =MAXIMUM CONCENTRATION OF U USED FOR PLOTTING        (MG/CM**3)
C.. VB    =VOLUME OF BULK LIQUID                                (CM**3)
C.. VL    =VOLUME OF LIQUID FILM                                (CM**3)
C.. F     =FLOW RATE OF INFLUENT                             (CM**3/SEC)
C.. FL    =FLOW RATE OF LIQUID FILM                          (CM**3/SEC)
C.. M     =NUMBER OF TIME STEPS
C.. N     =NUMBER OF TIME STEPS PER  SHAFT REVOLUTION
C..
C..          ********** INPUT FORMAT **********
C..            (TEN SPACES FOR EACH INPUT)
C..          UB        SO
C..          HE        P
C..          F         VB        AREA
C..          OMEGA     R         DEPTH
C..          PER       M
C..
      DIMENSION A(51),B(51),C(51),D(51),U(51),X(51),TS(51),W(51),
     1          Y(51),Z(51)
      DIMENSION AS(51),BS(51),CS(51),DS(51),S(51),TU(51),G(51),
     1          YS(51),ZS(51)
      COMMON /NAME1/ TLF,DIFF1,HE,P,UB /NAME2/ UMAX,SMAX,JUMP
      COMMON /NAME3/ ALPHA,OMEGA,ANGLE,THETA
C.. SPECIFIY THE CONSTANTS
      ITERC=0
      IBEGIN=1
      JUMP  =1
      EPS1  =0.01
      EPS2  =0.01
      SKL   =0.01
```

```
      DIFF1 =0.00005
      DIFF2 =0.0000064
      BMU   =0.000050
      BMICRO=20.
      STOIC =1.6
      RMS   =0.5
      B1    =0.001
      B2    =0.20
      THICK =0.005
      TLF   =0.01
C.. READ THE DATA
1     READ (5,1070,END=990) UB,SO,HE,P,F,VB,AREA,OMEGA,R,DEPTH,PER,
     1FINTIM,PLTIME,DX,THICK,PN
1070  FORMAT (3(7F10.0,/))
C..   CALCULATE DX,DT,FINTIM
      N=IFIX(PN+0.5)
      IF (N.LT.2) N=2
      DT=60./N/OMEGA
C.. GRID POINTS
      IDX=IFIX(THICK/DX+0.5)
      DX=THICK/FLOAT(IDX)
      IR=IFIX(THICK/DX)+1
C.. SPECIFY THE MAXIMUM VALUES OF S AND U FOR PLOTTING
      UMAX=0.0085
      SMAX=SO
C..  CALCULATE LIQUID FILM FLOW RATE
      FL=OMEGA/60.*AREA*TLF
C..  CALCULATE THE ANGLE OF SUBMERGENCE
      ALPHA=2.*ARCOS((DEPTH-R)/R)
C.. PRINT THE VALUES OF DX,DT,FINTIM
      WRITE (6,1010)
1010  FORMAT ('   SIMULATION BEGINNING ---- REX T. CHAN',//,' BIOLOGICAL
     1 DISC MODEL',/,' CRANK-NICOLSON SUBSTRATE REMOVAL RATE',/,
     2' VALUES OF TIME PARAMETERS **************************')
      WRITE (6,1020) DX,DT,FINTIM,PLTIME
1020  FORMAT (' DX=',F10.5,5X,'DT=',F10.5,5X,'FINTIM=',F10.0,5X,
     1'PLTIME=',F8.0)
      WRITE (6,1050) IR
1050  FORMAT (' IR= NUMBER OF GRID POINTS=',I3,//)
C.. WRITE OUT THE OTHER USEFUL INFORMATION
      WRITE (6,1055)
1055  FORMAT (' ******** FOR INFORMATION CONCERNING WITH PARAMETERS ****
     1****',/,' ******** NOMENCLATURE,DIMENSIONS AND VALUES SEE     ****
     2****',/,' ******** BEGINNING OF PROGRAM                       ****
     3****',//)
      WRITE (6,1060) UB,UMAX,SO,SMAX,F,FL,VB,AREA,HE,P,OMEGA,R,DEPTH,PER
     1,PN,THICK
1060  FORMAT (' INPUT PARAMETERS **************************',/,' UB=',5X
     1,F13.5,/,' UMAX=',3X,F13.5,/,' SO=',5X,F13.5,/,' SMAX=',3X,F13.5,/
     2,' F=',6X,F13.5,/,' FL=',5X,F13.5,/,' VB=',5X,F13.5,/,' AREA=',3X,
     3F13.5,/,' HE=',5X,F13.5,/,' P=',6X,F13.5,/,' OMEGA=',2X,F13.5,/,'
     4R=',6X,F13.5,/,' DEPTH=',2X,F13.5,/,' PER=',4X,F13.5,/,
```

```
     5' NUMBER OF TIME STEPS PER REV=',F4.0,/,' BIOFILM THICKNESS=',
     6F12.5)
      IR1=IR-1
C.. INITIALIZE THE ARRAYS,COUNTERS AND TIME
      PTIME=0.
      IFLAG=0
      INUM=6
      TIME=0.
      ANGLE=0.
      THETA=0.
      DO 10 I=1,IR
      A(I)=0.
      B(I)=0.
      C(I)=0.
      D(I)=0.
      AS(I)=0.
      BS(I)=0.
      CS(I)=0.
      DS(I)=0.
      G(I)=0.
      W(I)=0.
      Y(I)=0.
      YS(I)=0.
      Z(I)=0.
      ZS(I)=0.
      S(I)=BOUND2(-1.0)
      U(I)=BOUND2(-1.0)
      TU(I)=BOUND2(-1.0)
10    TS(I)=BOUND2(-1.0)
C.. CALCULATE THE X ARRAYS
      X(1)=0.
      DO 20 I=2,IR
20    X(I)=FLOAT(I-1)*DX
C.. INSERT THE BOUNDARY AND INITIAL  CONDITIONS FOR TIME=0
      TU(1)=UB
      TS(1)=0.01
      SL=0.01
      SB=0.01
      SBB=0.01
40    CONTINUE
C.. EXCHANGE THE OLD VALUES FOR THE NEW ONES
      DO 50 I=1,IR
      U(I)=TU(I)
50    S(I)=TS(I)
C.. CHECK TO SEE IF IT IS TIME TO PRINT
      IF (TIME-PTIME) 70,60,60
60    CALL DDUMP (TIME,U,S,X,IR,IFLAG,ITERC)
C.. IFLAG COUNTS PRINTS INTERVALS
      IF(IFLAG-INUM) 65,65,64
64    IFLAG=0
      PTIME=PTIME+PLTIME
      GOTO 70
```

```
65    IFLAG=IFLAG+1
70     ITERC=0
      IF (TIME-FINTIM) 80,900,900
80    IF (THETA.LT.3.7) GO TO 85
85    GO TO (90,100), IBEGIN
C.. DECREASE DT FOR THE FIRST TIME STEP
90    IBEGIN=2
      DDT=DT
      DT=0.2
      GO TO 139
100   DELTA=ALPHA-THETA
      IF (THETA.GT.(ALPHA-0.05).AND.(6.2832-THETA).GT.0.05) GO TO 130
      IF (DELTA.GT.0.05.AND.DELTA.LT.6.2832/N) GO TO 120
110   DT=DDT
      GO TO 139
120   DT=DELTA/(OMEGA*6.2832/60.)-0.01
      GO TO 139
130   DT=(6.2832-ALPHA)/IFIX(FLOAT(N)/2.+0.5)/(OMEGA*6.2832/60.)
139   INUM=60./(OMEGA*DT)
C.. UPDATE TIME
140   TIME=TIME+DT
C.. CALCULATE THE CONSTANTS
      C1=(DX)**2.
      C2=DIFF2*DT
      C3=2.*C1/C2
      C4=C1/DIFF2
      C5=BMU*BMICRO/RMS
      C6=DIFF1*DT
      C7=2.*C1/C6
      C8=C1/DIFF1
      C9=C5*STOIC
C.. PROJECT THE VALUE OF S AT THE NEW TIME LEVEL
      DO 150 I=2,IR
150   YS(I)=C5*DT/(B1+U(I))/(B2+S(I))*U(I)
C.. MIDDLE EQUATION
      DO 170 I=2,IR1
170   G(I)=(S(I+1)+S(I-1))*C2/C1+S(I)*(1.-2.*C2/C1-YS(I))
C.. LAST EQUATION
      I=IR
      G(I)=S(I)*(1.-C2/C1-YS(I))+S(I-1)*C2/C1
C.. PROJECT THE VALUES OF U AT THE NEW TIME LEVEL
      DO 200 I=2,IR
200   Y(I)=C9*DT/(B1+U(I))/(B2+S(I))*S(I)
C.. MIDDLE EQUATIONS
      DO 250 I=2,IR1
250   W(I)=(U(I+1)+U(I-1))*C6/C1+U(I)*(1.-2.*C6/C1-Y(I))
C.. LAST EQUATION
      I=IR
      W(I)=U(I)*(1.-C6/C1-Y(I))+U(I-1)*C6/C1
C.. THETA INDICATES IF THE DISC IS SUBMERGED OR EXPOSED TO THE AIR
      ANGLE=OMEGA/60.*6.2832*TIME
      THETA=ANGLE-FLOAT(IFIX(ANGLE/6.2832))*6.2832
```

```
      IF (THETA.GT.ALPHA) GO TO 400
C.. THIS  SECTION OF THE PROGRAM SOLVES THE BULK FLUID
C.. EQUATIONS BY THE MODIFIED EULER TECHNIQUE
C.. CALCULATE THE CHANGE OF SUBSTRATE CONCENTRATION IN THE LIQUID FILM
      SLL=SL
      IF (THETA.GT.6.2832/N) GO TO 320
      SLT=SB
      GO TO 340
320   SLT=SL
340   SLG=SLT-S(2)
C.. SLP IS TRIAL VALUE OF THE HALF TIME STEP
      SLP=SLT-(SKL/TLF*SLG*DT)
      TS(1)=SLP
      GO TO 500
C.. SL IS THE FINAL VALUE FOR THE FULL TIME STEP
350   SL=SLT-0.5*SKL/TLF*DT*(SLG+SLP-S(2))
      TS(1)=SL
      GO TO 500
C.. CALCULATE THE CHANGE OF SUBSTRATE CONCENTRATION IN THE BULK FLUID
400   SBB=SB
      E=F+FL+SKL*PER*AREA
      SBG=F*SO+FL*SL+SKL*PER*AREA*S(2)-SB*E
C.. SBP IS THE TRIAL VALUE FOR THE HALF TIME STEP
      SBP=SB+DT/VB*SBG
      TS(1)=SBP
      GO TO 500
C.. SB IS THE FINAL VALUE FOR THE BULK FLUID CONCENTRATION
420   SB=SB+0.5*DT/VB*(SBG+F*SO+FL*SL+SKL*PER*AREA*S(2)-SBP*E)
      TS(1)=SB
C.. MASS BALANCE ON S
500   DO 550 I=2,IR
550   ZS(I)=W(I)/(B2+G(I))/(B1+W(I))*C4*C5
C.. ENTRANCE BOUNDARY CONDITION
      I=2
      AS(I)=0.
      BS(I)=C3+2.+ZS(I)
      CS(I)=-1.
      DS(I)=S(I)*(C3-2.-ZS(I))+S(I+1)+S(I-1)+TS(1)
C.. MIDDLE EQUATIONS
      DO 570 I=3,IR1
      AS(I)=-1.
      BS(I)=C3+2.+ZS(I)
      CS(I)=-1.
570   DS(I)=S(I+1)+S(I)*(C3-2.-ZS(I))+S(I-1)
C.. EXIT BOUNDARY CONDITION
      I=IR
      AS(I)=-2.
      BS(I)=C3+2.+ZS(I)
      CS(I)=0.
      DS(I)=S(I)*(C3-2.-ZS(I))+2.*S(I-1)
C.. CALL THE THOMAS ALGORITHM TO SOLVE FOR S
      CALL TA (AS,BS,CS,DS,TS,IR)
```

```
C.. MASS BALANCE ON U
C.. INSERT THE BOUNDARY VALUE FOR U
      TU(1)=BOUND1(TIME)*UB+BOUND2(TIME)*UL(TIME)
      DO 600 I=2,IR
600   Z(I)=TS(I)/(B2+TS(I))/(B1+W(I))*C8*C9
C.. ENTRANCE BOUNDARY CONDITION
      I=2
      A(I)=0.
      B(I)=C7+2.+Z(I)
      C(I)=-1.
      D(I)=U(I)*(C7-2.-Z(I))+U(I+1)+U(I-1)+TU(1)
C.. MIDDLE EQUATIONS
      DO 650 I=3,IR1
      A(I)=-1.
      B(I)=C7+2.+Z(I)
      C(I)=-1.
650   D(I)=U(I+1)+U(I)*(C7-2.-Z(I))+U(I-1)
C.. EXIT BOUNDARY CONDITION
      I=IR
      A(I)=-2.
      B(I)=C7+2.+Z(I)
      C(I)=0.
      D(I)=U(I)*(C7-2.-Z(I))+2.*U(I-1)
C.. CALL THE THOMAS ALGORITHM TO SOLVE FOR U
      CALL TA (A,B,C,D,TU,IR)
C.. CHECK FOR CONVERGENCE
      IF (BMU) 990,40,700
700   CONTINUE
      TEST1=0.
      TEST2=0.
      DO 750 I=2,IR
      TEST1=TEST1+ABS(TU(I)-W(I))
750   TEST2=TEST2+ABS(TS(I)-G(I))
      IF (TEST1.GT.EPS1.OR.TEST2.GT.EPS2) GO TO 800
      IF (TS(1).EQ.SL.OR.TS(1).EQ.SLP) GO TO 770
      IF (SB.EQ.SBB) GO TO 420
      GO TO 40
770   IF (SL.EQ.SLL) GO TO 350
      GO TO 40
C.. AVERAGE AND ITERATE
800   DO 850 I=2,IR
      G(I)=TS(I)
850   W(I)=TU(I)
      ITERC=ITERC+1
      GO TO 500
900   CONTINUE
      GO TO 1
990   STOP
      END
      SUBROUTINE DDUMP(TIME,U,S,X,IR,IFLAG,ITERC)
C.. THIS SUBROUTINE DOES THE PRINTING AND PRINT PLOTING
C.. JUMP=1 SIGNALS THE FIRST TIME THROUGH THE SUBROUTINE ON THE FIRST
```

```
C.. TIME THROUGH A HEADING IS PRINTED AND CERTAIN CONSTANTS ARE
C.. CALCULATED
      INTEGER LINE(61),BLANK,STAR,SLASH
      DIMENSION U(1),S(1),X(1)
      COMMON /NAME2/ UMAX,SMAX,JUMP /NAME3/ ALPHA,OMEGA,ANGLE,THETA
      DATA BLANK/1H /,STAR/1H1/,SLASH/1H|/,IAS/1H2/
      GO TO (10,35),JUMP
10    JUMP=2
      P1=60./UMAX
      P2=60./SMAX
      WRITE(6,1000)
1000  FORMAT(1X,26HFINITE DIFFERENCE SOLUTION,//,13X,4HTIME,8X,
     18HDISTANCE,2X,14HCONCENTRATION1,2X,14HCONCENTRATION2,/,40X,8H(OXYG
     2EN),7X,11H(SUBSTRATE))
      K=IR/20
      IF(K.LT.1) K=1
      DO 20 I=1,61
20    LINE(I)=BLANK
35    WRITE (6,1050) THETA,IFLAG,ITERC
1050  FORMAT (////,'  LOCATION OF SIMULATION POINT=',2X,F10.5,1X,'RADIAN
     1S',5X,'PRINT NUMBER=',I3,5X,'ITERC=',I4,//)
      DO 40 I=1,IR,K
      LINE(1)=SLASH
      LINE(61)=SLASH
      S1=S(I)
      U1=U(I)
      INDEX1=IFIX(ABS(U1*P1+0.5))+1
      INDEX2=IFIX(ABS(S1*P2+0.5))+1
      IF(INDEX1.GT.61) INDEX1=61
      IF(INDEX1.LT.1) INDEX1=1
      IF(INDEX2.GT.61) INDEX2=61
      IF(INDEX2.LT.1) INDEX2=1
      LINE(INDEX1)=STAR
      LINE(INDEX2)=IAS
      X1=X(I)
      WRITE (6,1020) TIME,X1,U1,S1,LINE
1020  FORMAT (4E16.5,5X,61A1)
      LINE(INDEX2)=BLANK
40    LINE(INDEX1)=BLANK
50    RETURN
      END
      SUBROUTINE TA(A,B,C,D,Z,IR)
C.. THIS SUBROUTINE SOLVES THE SIMULTANEOUS EQUATIONS BY THE THOMAS
C.. ALGORITHM
      DIMENSION A(1),B(1),C(1),D(1),Z(1),BETA(51),GAMA(51)
      IR1=IR-1
      IR2=IR-2
C..PERFORM THE FORWARD CALCULATION OF THE THOMAS METHOD
C..FIRST EQUATION
      I=2
      BETA(I)=B(I)
      GAMA(I)=D(I)/B(I)
```

```
C..INTERIOR EQUATIONS
      DO 10 I=3,IR
      BETA(I)=B(I)-A(I)*C(I-1)/BETA(I-1)
      IF (D(I)-A(I)*GAMA(I-1).LE.1E-25) GO TO 300
      GAMA(I)=(D(I)-A(I)*GAMA(I-1))/BETA(I)
      GO TO 10
300   GAMA(I)=0.0
10    CONTINUE
C..PERFORM THE BACKWARDS CALCULATIONS
C..LAST EQUATION
      Z(IR)=GAMA(IR)
C.. INTERIOR AND FIRST EQUATIONS
      DO 20 I=1,IR2
      J=IR-I
20    Z(J)=GAMA(J)-C(J)*Z(J+1)/BETA(J)
      RETURN
      END
      FUNCTION BOUND1(TIME)
C.. THIS FUNCTION SETS THE ENTRANCE BOUNDARY CONDITION FOR U AND S
C.. WHEN ALPHA<ANGLE<=360(6.2832 RADIANS)
C.. NEGATIVE VALUES OF TIME INDICATE THAT THE INITIAL VALUE IS REQUESTED
      COMMON /NAME3/ ALPHA,OMEGA,ANGLE,THETA
      IF (TIME) 20,20,15
15    IF (THETA.GT.ALPHA) GO TO 10
      GO TO 20
10    BOUND1=1.
      RETURN
20    BOUND1=0.
      RETURN
      END
      FUNCTION BOUND2(TIME)
C.. THIS FUNCTION SETS THE ENTRANCE BOUNDARY CONDITION FOR U AND S
C.. WHEN ALPHA>=ANGLE>0
C.. NEGATIVE VALUES OF TIME INDICATE THAT THE INITIAL VALUE IS REQUESTED
      COMMON /NAME3/ ALPHA,OMEGA,ANGLE,THETA
      IF (TIME) 20,10,15
15    IF (THETA.LE.ALPHA) GO TO 10
      GO TO 20
10    BOUND2=1.
      RETURN
20    BOUND 2=0.
      RETURN
      END
      FUNCTION UL(TIME)
C.. THIS FUNCTION CALCULATES THE OXYGEN CONCENTRATION IN THE LIQUID
C.. FILM WHEN EXPOSED TO AIR
      COMMON /NAME1/ TLF,DIFF1,HE,P,UB /NAME3/ ALPHA,OMEGA,ANGLE,THETA
      IF (THETA.GT.ALPHA) GO TO 20
      TIME1=THETA/OMEGA*60./6.2832
      SUM=0.
      DO 10 M=1,10
      N=M-1
```

```
      X=(2.*N+1.)*TLF/2./SQRT(DIFF1*TIME1)
      A=2.*(-1)**N*ERFC(X)
10    SUM=SUM+A
      US=1777.8*P/HE
      UL=(US-UB)*SUM+UB
      RETURN
20    UL=UB
      RETURN
      END
      FUNCTION ERFC( X )
      DIMENSION B(5)
      DATA B /0.319382,-0.356564,1.781478,-1.821256,1.330274/
      DATA R,SQRTPI,SQRT2 /0.231642,1.772454,1.414214/
      IF(X.GT.7.5)GO TO 20
      Y = SQRT2*X
      FY =  EXP(-Y**2/2.)/SQRT2/SQRTPI
      T = 1./(1.+R*Y)
      Q = B(4) + B(5)*T
      DO 10 I=1,3
   10 Q = B(4-I) + T*Q
      ERFC = 2.*FY*Q*T
      RETURN
   20 ERFC = 0.
      RETURN
      END
```

APPENDIX B

EXPERIMENTAL DATA

BOD5

DAYS*	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4	CLARIFIED EFFLUENT
1	51.0	24.0	23.0	18.7	11.7	10.0
7	185.0	29.5	19.1	14.8	12.6	6.8
9	129.2	36.1	29.9	22.5	13.0	7.9
14	180.1	34.6	21.3	15.1	7.0	4.2
16	147.5	47.6	28.6	18.2	13.5	4.3
21	176.1	33.7	22.4	12.9	11.4	3.3
23	119.9	28.3	23.8	13.9	12.2	1.5
28	148.6	43.4	23.7	18.3	8.9	2.2
30	240.0	46.1	20.9	12.9	8.0	1.6
34	93.4	31.3	19.2	13.0	7.0	1.3
42	114.7	16.3	8.4	5.5	4.1	1.1

COD

DAYS	INFLUENT	EFFLUENT		
		TOTAL	CLARIFIED	SOLUBLE
5	350.0	89.3	46.3	17.9
7	387.0	79.4	25.8	13.9
9	231.2	86.8	37.7	24.5
14	402.0	57.0	38.0	24.7
16	300.0	86.8	37.7	26.4
21	429.1	39.4	27.6	19.7
23	261.7	85.2	36.5	24.3
28	310.1	59.1	39.4	23.6
30	396.3	51.6	27.5	20.5
34	159.5	38.9	3.9	3.0
42	264.7	33.1	13.6	13.6

*Days are the no. of days after the first day of data collection

All concentrations are in mg/l unless specified.

NITRATE - NITROGEN

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
1	.65	29.49	32.00	32.00	32.00
4	.20	4.97	5.53	5.65	5.76
5	.34	22.83	30.01	34.30	36.83
6	.20	37.51	42.99	42.99	41.01
8	.86	28.70	37.53	38.07	41.30
9	2.08	39.47	42.27	39.47	38.50
12	.56	25.63	30.10	31.77	34.44
13	.88	33.87	34.73	33.87	33.33
14	3.16	32.90	37.10	37.10	38.07
15	3.30	39.20	43.40	42.00	40.60
16	2.60	42.00	49.70	49.00	46.47
21	.72	43.13	42.70	42.70	42.70
22	.68	36.13	40.17	41.57	43.67
23	.86	38.03	42.70	42.70	43.67
26	.59	21.84	26.60	31.50	34.30
27	.65	36.40	39.90	37.53	34.73
28	.75	45.07	50.13	49.70	47.33
29	.54	37.10	43.40	49.00	51.80
30	.50	39.90	45.50	46.20	48.30
33	1.13	50.40	51.10	51.10	51.10
34	.77	51.10	55.30	54.60	53.90
35	.63	39.47	44.80	49.70	53.76
40	.88	24.93	24.93	28.00	30.80
41	.65	46.90	49.70	42.70	36.13
42	.77	38.64	43.40	46.90	46.90

AMMONIA - NITROGEN

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
1	15.98	4.00	.20	.20	.20
4	31.71	4.94	3.62	2.26	2.02
5	18.45	.99	.20	.20	.20
6	39.04	2.85	1.62	1.55	1.40
8	24.54	.74	.20	.20	.20
9	42.74	1.32	.20	.20	.20
12	25.20	1.40	.20	.20	.20
13	35.25	1.33	.20	.20	.20
14	32.94	1.73	.20	.20	.20

CONTINUED

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
15	44.14	2.64	.20	.20	.20
16	43.65	4.20	.49	.20	.20
21	37.55	9.55	1.40	.20	.20
22	34.01	.99	.20	.20	.20
23	38.05	1.15	.20	.20	.20
26	20.75	.58	.20	.20	.20
27	45.54	7.41	4.04	1.07	.20
28	44.80	1.24	.20	.20	.20
29	30.39	.58	.20	.20	.20
30	36.81	1.04	.20	.20	.20
33	46.28	1.03	.20	.20	.20
34	49.00	1.82	.20	.20	.20
35	33.60	2.55	.33	.20	.20
40	20.01	1.56	.20	.20	.20
41	32.45	.96	.20	.20	.20
42	28.00	2.59	.38	.20	.20

TOTAL SUSPENDED SOLIDS

DAYS	INFLUENT	EFFLUENT	
		TOTAL	CLARIFIED
5	49.7	51.8	23.5
16	90.0	55.5	2.3
21	93.9	26.5	6.1
22	78.0	40.8	4.5
28	88.5	29.5	2.1
29	70.9	27.2	4.3
30	80.6	25.9	2.9
34	53.7	21.5	1.5
42	54.1	9.6	1.1

pH

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
1	8.45	7.15	6.68	5.72	5.45
4	7.10	6.70	5.80	5.30	5.28
5	7.65	6.85	6.30	5.55	5.25
6	7.40	6.80	6.20	5.60	5.10
8	7.30	7.77	8.30	8.40	8.39
9	8.30	7.40	7.85	8.05	8.10
12	7.84	7.63	8.09	8.19	8.21
13	8.40	7.61	8.03	8.25	8.19
14	7.62	7.19	7.28	7.76	7.92
15	7.52	6.78	6.71	7.03	7.30
16	8.41	6.88	6.67	6.97	6.99
21	8.10	7.53	7.35	7.62	7.82
22	6.96	7.23	7.61	7.84	7.84
23	8.12	7.31	7.75	7.88	7.84
26	7.25	7.31	7.55	7.66	7.71
27	8.19	6.80	6.24	6.26	7.36
28	7.64	8.17	8.16	7.89	7.84
29	7.84	7.36	7.89	8.09	8.10
30	8.28	6.50	6.70	7.53	7.84
33	8.28	6.55	7.27	7.61	7.55
34	8.61	7.08	7.78	7.89	7.75
35	7.97	6.44	6.75	7.27	7.56
40	8.17	7.42	8.09	8.14	8.06
41	8.36	6.58	7.74	8.08	8.14
42	7.88	6.17	6.24	7.15	7.61

NITRITE - NITROGEN

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
7	.50	6.56	2.84	.55	.10
10	.66	4.88	1.48	.15	.02

DISSOLVED OXYGEN

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
16	1.42	6.38	7.75	7.82	7.82
30	.40	6.00	8.80	9.20	8.22
42	.40	5.65	8.50	9.05	9.00

TEMPERATURE (in Degree Centigrade)

DAYS	INFLUENT	STAGE 1	STAGE 2	STAGE 3	STAGE 4
30	23.0	18.0	17.0	17.0	17.0
42	24.0	19.2	18.2	17.5	17.5

PHOSPHATE

DAYS 1 INFLUENT 5.68

BIOFILM THICKNESS

DAYS 43 AVERAGE THICKNESS 150 microns

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