EFFECT OF CELL RESIDENCE TIME ON NITRIFICATION WITH A ROTATING BIOLOGICAL CONTACTOR SYSTEM

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EFFECT OF CELL RESIDENCE TIME ON NITRIFICATION

WITH

A ROTATING BIOLOGICAL CONTACTOR SYSTEM

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SUMMARY

The effect of cell residence time (θ_c) on biological nitrification of a simulated secondary wastewater effluent was evaluated using a bench-scale rotating biological contactor (RBC) system. θ_c values were controlled by periodic scraping of attached biomass from disc surfaces and by continuous circulation of RBC mixed liquor to remove sloughed biomass. θ_c values near that for washout of nitrifying microorganisms were investigated. The effects of an influent organically-bound nitrogen and influent wastewater organic strength on nitrification were examined.

The critical θ_c value for washout of nitrifying microorganisms was approximately 1.5 d. This value was similar to results reported for pure cultures of Nitrosomonas and Nitrobacter and for mixed cultures in wastewater treatment systems. Hydrolysis of the organically-bound nitrogen was not a limiting step in nitrification during the experiment using glycine as the sole nitrogen source. Organic loadings as high as 13.9 g COD/d·m^2 had no effect on nitrification at influent ammonia-nitrogen and hydraulic loadings of 1.4 g N/d·m^2, respectively.

The results indicated that RBC systems can be utilized for nitrification of domestic wastewaters with or without prior biological treatment.

CHAPTER I

INTRODUCTION

Stringent effluent requirements on domestic and industrial discharges are proposed for 1983 by EPA. The organic matter and ammonia, as well as other substances permitted in wastewater discharges, may be decreased from the present requirements. Many wastewater treatment facilities are meeting the present effluent limits with secondary biological treatment. However, additional treatment will be needed by many of these facilities to remove the organic matter and ammonia required to meet the future limitations. One alternative is to treat the present secondary effluent with a rotating biological contactor (RBC) system. RBC systems are capable of removing organic matter and ammonia. The systems are presently designed based on flow, influent organic concentration and the desired effluent organic concentration. These systems normally have cell residence times from three to more than 30 days. Cell residence time (θ_c) is the period of time a given cell or unit mass of biological film is retained within the RBC system. Antonie (1976) stated that θ_c cannot be controlled; therefore, it is not considered to be a

controlling parameter. However, studying the RBC process with respect to θ_c will provide useful process information.

The objective of this research project was of dual purpose: 1. investigate the effectiveness of a bench-scale RBC system, treating a synthetic domestic secondary effluent, in removing organic matter and achieving nitrification; and 2. determine the effects of θ_c on nitrification. In addition, the effect of influent organic matter concentration was investigated at low θ_c values (2 to 4 d). Also, the hydrolysis of a nitrogenous organic compound and the subsequent oxidation of the ammonia produced was investigated with the RBC system. The data collected by Cruz (1977) with the same RBC system were analyzed in conjunction with the data collected in the present study.

CHAPTER II

LITERATURE REVIEW

General Description

The rotating biological contactor (RBC) process consists of an inert rotating media and a biological population. The population includes organisms attached to the media and suspended organisms in the mixed liquor. The mass of organisms attached to the media are contained in a layer referred to as a biofilm. The media, to which the biofilm is attached, is supported by a rotating horizontal shaft. The media is partially submerged in a basin containing wastewater such that any section of the media is alternately passing through the wastewater and the atmosphere.

The support media in a full-scale RBC system is made of expanded polystyrene (Antonie & Van Aacken, 1971; Steels, 1974) and can be either smooth (Hartmann, 1960; Popel, 1964; Steels, 1974) or corrugated (Antonie, 1976) discs. The media specific surface area, i.e. surface area per unit volume of media, varies from 15 to 37 ft^2/ft^3 (Antonie, 1976; Stover and Kincannon, 1976). The media is typically submerged approximately 40% into the wastewater (Antonie, 1976;

Birks and Hynek, 1971) and is rotated at speeds from one to five rpm (Antonie, 1976; Birks and Hynek, 1971; Steels, 1974). Antonie (1976) reported that the optimum peripheral velocity for a domestic wastewater was approximately 60 ft/min, e.g. a 5.8 ft diameter disc would be rotating at 3.3 rpm. The diameter of the media or discs range from 3 to 12 ft (Antonie, 1976; Birks and Hynek, 1971; Chittenden and Wells, 1971; Steels, 1974). A typical shaft, to which the media is attached, ranges from 6 to 25 ft in length (Antonie, 1976; Steels, 1974).

The physical structure of a RBC process has several advantages over the activated sludge and trickling filter processes. The RBC process has low power requirements because the buoyancy of the plastic discs offsets the weight of the discs, biomass and the support structure so that the shaft structure has a minimal resultant downward force (Antonie and Van Aacken, 1971). For an activated sludge process the power requirements will be more than double the requirements for a RBC process treating the same waste to the same level of treatment (Antonie, 1976). For example, a one mgd plant with an inlet BOD_5 concentration of 150 mg/1 and a required effluent BOD_5 concentration of 15 mg/1 would require 55 HP/mgd for an activated sludge process and only 16 HP/mgd for a RBC pro-

cess (Antonie, 1976). The headloss through a RBC process is usually less than 1 ft (Antonie and Van Aacken, 1971) which is considerably less than the 5 to 10 ft of headloss through a trickling filter (Metcalf and Eddy, 1972). The nuisances associated with trickling filters such as clogging, Psychoda flies and objectionable odors, are absent (Antonie and Van Aacken, 1971).

Biofilm Theory

Biofilms have been studied by numerous researchers (Kornegay and Andrews, 1969; Tomlinson and Snaddon, 1966; Hoehn, 1970; Grieves, 1972; Atkinson et al., 1967) using various types of media such as discs, pipes and drums, and fixed plates. The results from these studies will be used to describe biofilms in RBC systems.

The organisms in a biofilm have been classified into two layers, i.e. an active film and an inactive film, with respect to the removal of soluble organic matter (Kornegay and Andrews, 1969; Sanders, 1964; Water Pollution Research, 1957). This concept is presented in Figure 1 (Kornegay and Andrews, 1969). The active film thickness, d, constitutes only a portion of the total film thickness, h (Kornegay and Andrews, 1969). Sanders (1966) found that as the thickness of a microbial film

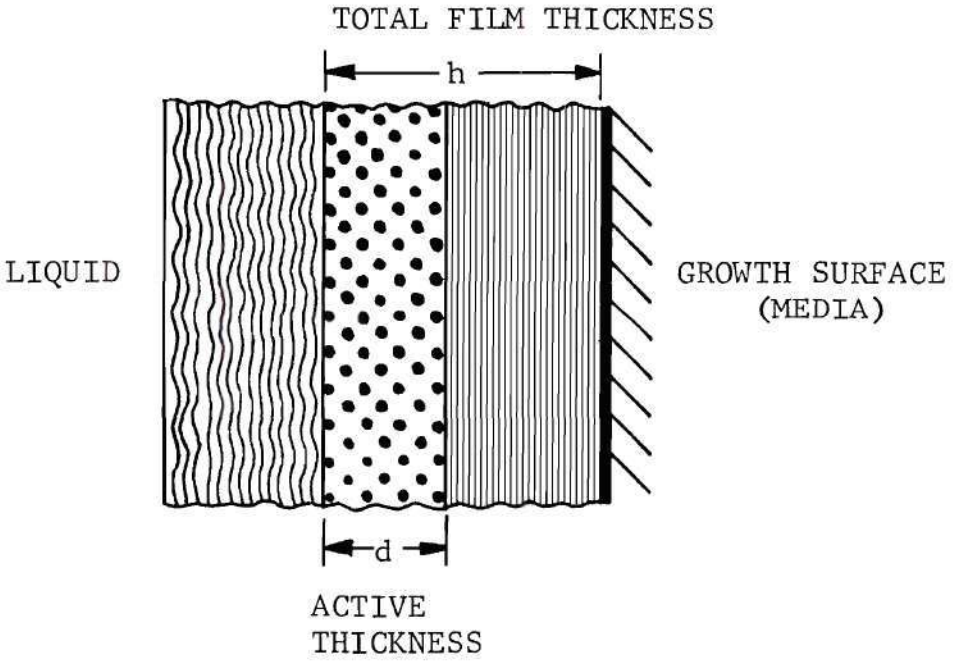

Figure 1. Cross-Section of a Biofilm

(Kornegay and Andrews, 1969)

increased to a critical film depth, substrate uptake rates increased. Beyond this critical film depth, no further increase in substrate uptake occurred. In addition, the existence of an active film has been confirmed by many others (Hoehn, 1970; Kornegay and Andrews, 1969; Eckenfelder, 1961; Wuhrmann, 1963; Tomlinson and Snaddon, 1966). Tomlinson and Snaddon (1966) suggested that the active portion of the film was the aerobic region. However, Kornegay and Andrews (1969) stated that active thickness and aerobic film thickness were not necessarily the same and that the active film thickness was a basic property of the particular substrate-microbe system. Atkinson and Fowler (1974) suggested that active film thickness for biofilms in fermentation applications was equal to the depth of penetration of the substrate, since uptake rates increased until the film thickness was equal to the penetration depth.

When the active depth of a biofilm reaches the thickness at which maximum substrate occurs and no additional growth in thickness of the biofilm improves the substrate removal, then this thickness is referred to as the "critical depth". Many researchers have observed this critical depth phenomena and their results are recorded in Table 1. The depths vary by approximately one order of magnitude. The difference in these

Table 1

Critical Depths of Microbial Films

Critical Depth (μm)	Type of System	Type of Substrate	Reference
65	rotating drum	glucose	Kornegay and Andrews, 1969
27-62	culture chamber and Plexiglas slide	synthetic substrate	Sanders, 1966
200	inclined rotating tube	domestic sewage	Tomlinson and Snaddon, 1966
120	inclined rotating tube	domestic sewage	Water Pollution Research, 1957
150	rotating cylinder	synthetic substrate	Hoehn, 1973

values may be the result of the environment to which the biofilms were subjected.

Howell and Atkinson (1976) presented five possible conditions that could occur in a biofilm. These are shown in Figure 2. The organisms in the biofilm could be substrate limited (Figure 2-A); oxygen limited (Figure 2-B); or limited by both oxygen and substrate (Figure 2-C). Oxygen or substrate could also become limiting to the lower layers of organisms as the biofilm thickness increased, respectively (Figure 2-D and E). The concentration of substrate and oxygen in the biofilm is controlled by two competing mechanisms, diffusion and metabolism (Lee et al., 1976). For oxygen or substrate the utilization rate may be controlled by the rate of transport to the reaction site or by the kinetics of biochemical reactions.

Lee et al. (1976) stated that the reaction kinetics between dissolved oxygen and organisms in the biofilm did not depend on the dissolved oxygen concentration; therefore, the reaction was zero order with respect to the oxygen concentration in the mixed liquor. Sanders (1966) indicated that the mass transfer of oxygen into the liquid phase may control oxygen utilization. Tomlinson and Snaddon (1966) also concluded that a higher oxygen concentration above the film would

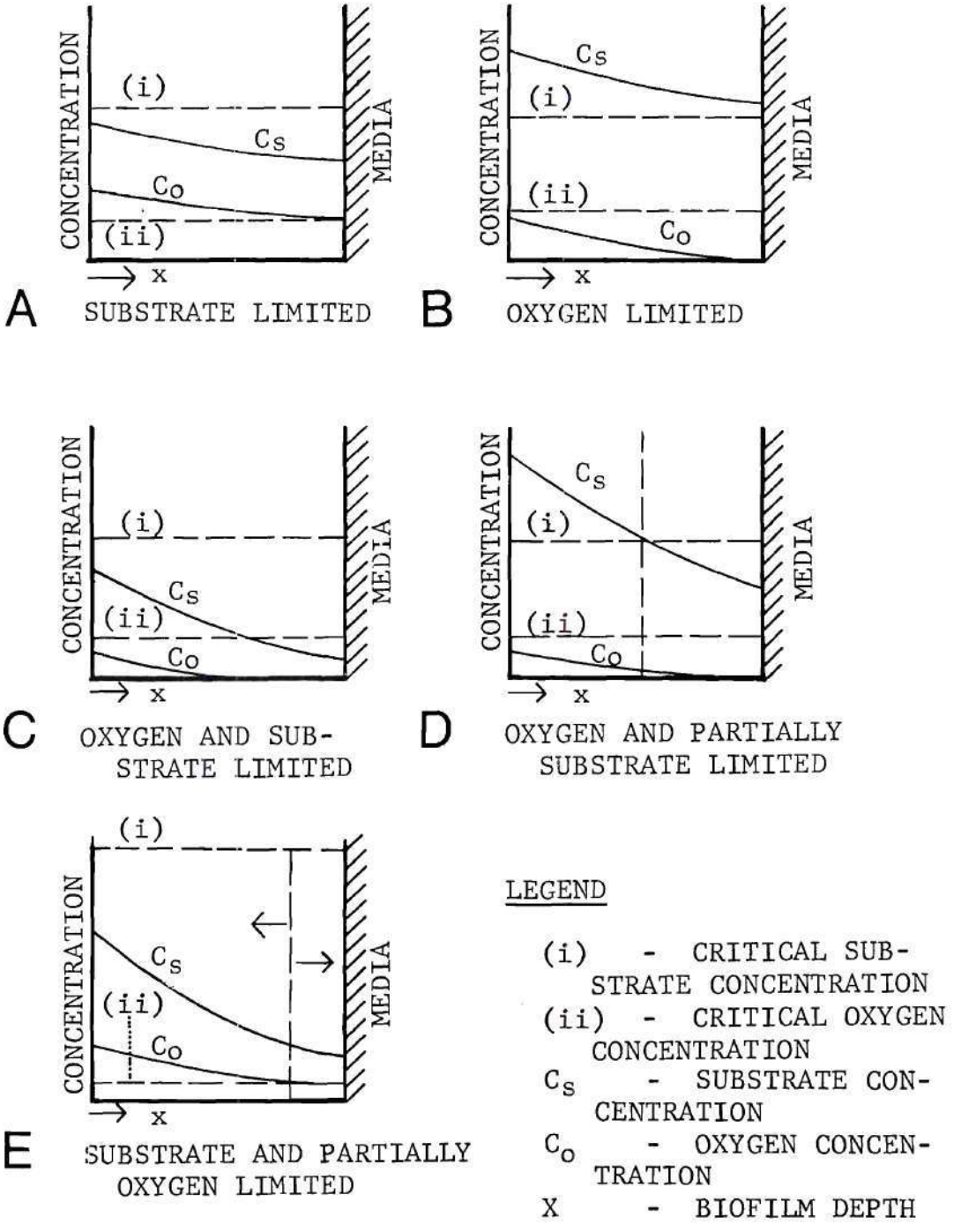

Figure 2. Possible Oxygen and Substrate and Substrate Concentration Profiles in a Biofilm (Howell and Atkinson, 1976)

increase diffusion of oxygen into a biofilm and increase oxygen utilization, i.e. first order uptake. Mehta et al. (1972) developed a performance equation for plastic media trickling filters based on the assumption that oxygen transport in the liquid phase limited the organic removal rate. The predicted organic removal capacities were in agreement with reported removal levels; therefore, supporting the assumption that liquid phase oxygen transport limited the removal of organic matter. Rahn and Richardson (1941) studied the oxygen requirements of pure cultures of bacteria growing in the presence of excess substrate and found that the oxygen requirements of the bacteria examined varied considerably. It was also noted that oxygen requirements increased with temperature.

The information on oxygen uptake presently available conflicts to such an extent that the true effect of the dissolved oxygen concentration on film activity is difficult to accurately assess. From the literature, oxygen utilization is determined by the species present (Rahn and Richardson, 1941), the diffusion rate of the oxygen from the aqueous phase into the biofilm, the metabolism by the organisms (Lee et al., 1976) and whether the substrate is limiting growth (Howell and Atkinson, 1976).

Williamson and McCarty (1976) stated that substrate removal by a biofilm from an aqueous phase required diffusion of all metabolic reactants into the biofilm, metabolism by the organisms, and diffusion of the metabolic products back through the biofilm into the aqueous phase. Metabolic uptake and diffusion rates of substrate and oxygen may control substrate removal (Lee et al., 1976).

As the organisms in the biofilm continue to remove substrate and subsequently multiply, the biofilm increases in depth. Howell and Atkinson (1976) stated that as a biofilm increased in thickness, sloughing began to increase due to deterioration of the adhesive properties of the lower layers. Tomlinson and Snaddon (1966) presumed that sloughing was caused by anaerobic breakdown in the film base. As a biofilm increases in depth, the concentrations of substrate and oxygen decrease in the base layers and these changes may affect organism metabolic systems which may then affect cellular adhesive properties. An increased thickness in a biofilm also increases the weight of the film which increases the shear stress on the base organisms. Sloughing occurs when the shearing forces become greater than the adhesive forces. After a section of biofilm has sloughed off the support media, the remaining attached organisms grow and produce a new

biofilm. The sloughing and rebuilding of a biofilm cycle has been observed by Bintanja et al. (1976), Hudson et al. (1976), Welch (1968) and Tomlinson and Snaddon (1966).

Heterotrophs

The organisms in a biofilm can generally be divided into two classes, heterotrophs and autotrophs (Metcalf and Eddy, 1972). Welch (1968) observed that the predominate heterotrophic organisms in a biofilm were bacteria and fungi. The fungi were the main support media for the bacteria. Aerobic heterotrophs use organic matter as electron donors and oxygen as electron acceptors; organic matter is also used for cell synthesis. Bacteria can be classified according to the temperature range in which they function best. Metcalf and Eddy (1972) stated that the classifications are cryophilic, mesophilic and thermophic with existence occurring in temperature ranges between -2°C and 30°C, 20°C and 45°C, and 45°C and 75°C, respectively. The optimum temperature ranges are between 12°C and 18°C, 25°C and 40°C, and 55°C to 65°C, respectively. The optimum pH was reported between 6.5 and 7.5. Clark et al. (1971) stated that bacteria require nitrogen and phosphorus; the requirements are based on the amount of BOD_5 utilized. The ratio of BOD_5 to nitrogen to phosphorus was

reported to be 100 to 5 to 1.

Metcalf and Eddy (1972) stated the optimum pH for most fungi species was 5.6 with an existence range between 2 and 9. Also, fungi required only half the nitrogen used by bacteria.

Autotrophs

The other class of organisms in a biofilm, autotrophs, are bacteria that use ammonium and nitrite ions as electron donors and oxygen as electron acceptors. The ability to oxidize inorganic nitrogen, i.e. nitrification, can be accomplished by a wide variety of heterotrophic bacteria as well as the autotrophic bacteria (Cutler and Crump, 1933; Fischer et al., 1956). However, the heterotrophic bacteria typically have a low nitrification rate compared to the autotrophic nitrifying bacteria of the Nitrobacteriaceace family. Painter (1970) concluded that nitrification in wastewater treatment processes was primarily due to the activity of the two genera, Nitrosomonas and Nitrobacter.

The autotrophic nitrifying bacteria derive energy released by the oxidation of an inorganic nitrogen substrate, ammonia and nitrite. Carbon dioxide or bicarbonate are the carbon sources for cellular synthesis. The stoichiometric reaction for oxidation of ammonium to nitrite by Nitrosomonas and nitrite to nitrate by Nitrobacter are, respectively:

$$NH_4^+ + \frac{3}{2}O_2 \xrightarrow{\text{Nitrosomonas}} 2H^+ + H_2O + NO_2^- \qquad (1)$$

$$NO_2^- + \tfrac{1}{2}O_2 \xrightarrow{\text{Nitrobacter}} NO_3^- \qquad (2)$$

Free energy changes have been reported for the first reaction to vary between 58 kcal/mole and 84 kcal/mole and for the second reaction to vary between 15.4 kcal/mole and 20.9 kcal/mole (Baas-Becking and Parks, 1927; Haug and McCarty, 1971; Lees, 1954; Nicholas, 1963; Painter, 1970). Kluyver and Donker (1926) postulated that there are two intermediates in the oxidation of ammonia to nitrite. Hofman and Lees (1953) report the first intermediate to be hydroxylamine (NH_2OH). The oxidation product of hydroxylamine is as yet undetermined. Aleem et al. (1965) showed that the oxidation of nitrite to nitrate can be separated into two coupled reactions; no intermediates have been isolated in this enzymatic oxidation.

Oxygen is utilized in the oxidation of both ammonium and nitrite as shown in equations (1) and (2). Assuming no nitrogen is used for synthesis, the stoichiometric quantities required are 3.43 mg oxygen/mg ammonium-nitrogen and 1.14 mg oxygen/mg nitrite-nitrogen, respectively. Gujer and Jenkins (1974) proposed the following equations to take into account

the nitrogen that is used for synthesis. The overall synthesis and oxidation reaction was:

$$NH_4^+ + 1.83O_2 + 1.98HCO_3^- \longrightarrow 0.021C_5H_7NO_2 + 1.041H_2O + 0.98NO_3^- + 1.88H_2CO_3 \quad (3)$$

Gujer and Jenkins (1974) assumed yields for Nitrosomonas and Nitrobacter of 0.15 mg cells/mg ammonium-nitrogen and 0.02 mg cells/mg nitrite-nitrogen, respectively. Also, the cell composition was assumed to be $C_5H_7NO_2$. The stiochiometric oxygen consumption based on equation (3) was 4.19 mg oxygen/mg ammonium-nitrogen. Wezernak and Gannon (1967) used BOD incubation bottles and obtained values of 3.22 mg oxygen/mg ammonium-nitrogen and 1.11 mg oxygen/mg nitrite-nitrogen with standard deviations of 0.06 and 0.01, respectively. Jeffrey and Morgan (1959) found oxygen uptake values in long term BOD tests for nitrification were within 2.5 percent of the theoretical values calculated with equations (1) and (2). The long term study minimized the effects of synthesis by endogenous respiration. The variations in the above results are probably from the bacteria being exposed to different environmental conditions.

The rate of nitrification has been reported to be affected by pH (Hall, 1973 ; Rimer and Woodward, 1972; Wild

et al., 1972; Huang and Hopson, 1974). The rate generally exhibits a plateau over an optimum pH range and decreases on either side of the plateau. Hall (1973), Rimer and Woodward (1972) and Wild et al. (1972) reported optimum pH ranges for nitrification in activated sludge were 7.0 to 9.4, 8.4 to 8.5 and 8.0 to 8.6, respectively. Huang and Hopson (1974) stated the optimum pH range for nitrification in an attached growth reactor was 8.4 to 8.8. The difference in the reported optimal pH ranges may be attributed to the pH at which each culture was acclimated. The ability of nitrifying bacteria to acclimate to a new pH has been observed by Haug and McCarty (1971) and Stankewich (1972). Haug and McCarty (1971) reported that for an attached growth system, complete acclimation to a pH of 6.0 was obtained after 10 days; the initial pH was between 7.0 and 8.5. Stankewich (1972) observed high degrees of nitrification between a pH of 5.8 and 6.0.

The nitrification reaction results in a release of hydrogen ions, as shown in equation (1), and may result in a decrease in pH. Without proper acclimation, low pH values can decrease the rate of nitrification. Therefore, the alkalinity of the wastewater is important. Using equation (3), 7.07 mg of alkalinity as $CaCO_3$ is destroyed per mg of ammonia-nitrogen oxidized. For suspended growth systems, Mulbarger (1971), Horstkotte

et al. (1974) and Newton and Wilson (1973) observed values of 6.4, 6.0 and 7.1 mg alkalinity as $CaCO_3$/mg ammonium-nitrogen, respectively. Gasser et al. (1974), Osborn (1965) and Haug and McCarty (1971) reported values of 6.5, 6.3 to 7.4 and 7.3 mg alkalinity as $CaCO_3$/mg ammonium-nitrogen, respectively, for attached growth cultures. The variance in these may be contributed to the environmental conditions to which the cultures were subjected.

Temperature is of fundamental importance to the growth of microorganisms since their metabolic rates are, to a large extent, dependent on it. The optimum temperature for nitrification has been reported to be 30°C for attached and suspended growth, with a decrease in nitrification rate if the temperature is decreased or increased (Buswell et al., 1954; Haug and McCarty, 1971; Balakrishnan et al., 1969; Haung and Hopson, 1974; Downing et al., 1964 and 1965). The above six references indicated that attached growths have a higher rate of nitrification at lower temperatures than suspended growth cultures; this phenomenon is due to nitrifiers being washed out of the suspended growth system at low growth rates but remaining in the attached growth system. However, Hall (1973) maintained full nitrification in a pilot scale activated sludge plant at temperatures as low as 9°C. For the upper

range of temperatures, Gibbs (1920) stated that exposure to 53°C to 55°C for ten minutes inactivated Nitrosomonas and an equal exposure to 56°C to 58°C inactivated Nitrobacter.

Oxygen can be a growth-limiting substance to nitrifiers Nagel and Haworth (1969) found that as dissolved oxygen increased from 0.25 mg/l to 2.0 mg/l, the rate of ammonia oxidation also increased. Oxygen was concluded to be limiting at low concentrations and appeared to follow the typical Monod (1942) relationship. Downing and Bayley (1961) and Jenkins (1969) reported that the lowest oxygen concentration at which nitrification could occur was 0.5 mg/l. Downing and Scragg (1958) reported the limiting concentration to be 0.3 mg/l. In three identically operated high-rate activated sludge pilot plants having 1, 4, and 7 mg/l of dissolved oxygen, Wuhrman (1963) noted that nitrification was only about 10 % complete at 1 mg/l while it was 90 % complete at the two higher concentrations. Downing et al. (1964) stated that if the concentration of dissolved oxygen exceeded 1 mg/l, nitrification would not be limited. In regard to nitrification at high dissolved oxygen concentrations, Haug and McCarty (1971) reported that 60 mg/l did not affect the growth rates of nitrifying bacteria. Okun (1949) noted that nitrification was not affected by dissolved oxygen concentra-

tions at 33 mg/l.

Carbon dioxide,is required for growth and is also the source of carbon for cellular synthesis. The ammonium and nitrite ions are the source of energy for Nitrosomonas and Nitrobacter, respectively. There are some nutrients that are essential for nitrifiers and others that are stimulatory. The rate of growth of a bacterial culture is a function of the concentration of the limiting nutrient or substrate. The growth characteristics of nitrifying bacteria have been studied (Downing et al., 1964; Lijklema, 1973) and appear to conform to an expression presented by Monod (1942):

$$\mu = \hat{\mu} \frac{S}{K_S + S} \tag{4}$$

where μ = specific growth rate (T^{-1}), $\hat{\mu}$ = maximum specific growth rate coefficient (T^{-1}), K_S = saturation coefficient (M/L^3) and S = limiting substrate concentration (M/L^3). Therefore, at low substrate concentrations the rate of growth is directly proportional to the substrate concentration, i.e., a first order reaction since $K_S >> S$. At high substrate concentrations, however, the growth rate approaches a maximum value and is independent of substrate concentration, i.e., a zero

order reaction since $K_s + S \cong S$. Monod (1942) originally developed equation (4) to fit the declining phase of bacterial growth for batch cultures; Monod also used this equation to model continuous cultures. Downing et al. (1964) and Lijklema (1973) also found equation (4) to fit continuous growth rates of nitrifiers.

Both μ and K_s are functions of temperature as shown by Knowles et al. (1965) for Nitrosomonas as:

$$\log_{10} \hat{\mu} = 0.0413\ (T) - 0.944 \qquad (5)$$

$$\log_{10} K_s = 0.051\ (T) - 1.158 \qquad (6)$$

and for Nitrobacter as:

$$\log_{10} \hat{\mu} = 0.0255\ (T) - 0.492 \qquad (7)$$

$$\log_{10} K_s = 0.063\ (T) - 1.149 \qquad (8)$$

where $\hat{\mu}$ is expressed in d^{-1}, K_s is expressed in mg nitrogen/l and T is expressed in ^{o}C. Poduska (1973) presented a comparison between heterotrophs, Nitrosomonas and Nitrobacter using yield coefficients, maximum specific growth rates and saturation coefficients. The values utilized were ranges of values obtained from other research (Hall, 1973; Downing, 1964; Lees and Simpson, 1957; Laudelot and Tichelen, 1960; Haug and

McCarty, 1971; Boon and Laudelot, 1962; Lees, 1952; Loveless and Painter, 1968; Buswell et al., 1954; Knowles et al., 1965; Gould and Lees, 1960; Melamed et al., 1970; Domey et al., 1971) and are recorded in Table 2. The low values of K_s for the nitrifying bacteria indicate that when the concentration of either ammonium or nitrite ions is greater than 2 to 3 mg/1, the growth rate will be at a maximum value for both species and therefore zero order with respect to the substrate concentration (Poduska, 1973).

Many biological systems exhibit growth rate characteristics similar to the Monod function over a wide range of substrate concentrations. However, most organisms also exhibit some form of substrate inhibition if the substrate concentration is raised to a sufficiently high level. Meyerhof (1916 and 1917) determined that the maximum respiration rates for Nitrosomonas and Nitrobacter occurred at approximately 110 mg ammonia-nitrogen/1 and 210 mg nitrite-nitrogen/1, respectively. Above these concentrations, the growth rates were reported to decrease. However, more recently Engel and Alexander (1958) reported no inhibition for Nitrosomonas in a pure culture with a pH of 8.0 and at a concentration of 640 mg ammonia-nitrogen/1. Anthonisen et al. (1976) stated that inhibition of nitrifiers was related to the concentration of free ammonia (NH_3) and nitrous acid (HNO_2). The

Table 2

Comparison of Stiochiometric and Kinetic Coefficients For Nitrifying and Heterotrophic Bacteria (Poduska, 1973)

	Heterotrophs	Nitrosomonas	Nitrobacter
Y (g/g)	0.4 - 0.6	0.03 - 0.1	0.02 - 0.08
$\hat{\mu}$ ($hours^{-1}$)	0.15 - 0.2	0.04 - 0.08	0.04
K_s (mg/l)	100 - 200*	0.18 - 1.0	0.25 - 1.0

*Apparent values for activated sludge systems.

ranges of free ammonia-nitrogen that began to inhibit Nitrosomonas and Nitrobacter were 10 to 150 mg/1 and 0.1 to 1.0 mg/1, respectively. The concentrations of free ammonia that inhibit Nitrosomonas are greater than those that inhibit Nitrobacter. Nitrous acid was observed to inhibit only Nitrobacter; the inhibition was initiated at concentrations between 0.22 and 2.8 mg/1. From the graphs presented by Anthonisen et al. (1976) a total ammonia-nitrogen concentration of 1.5 mg/1 at a pH of 8.3 would have a free ammonia-nitrogen fraction of 0.1 mg/1, the concentration reported to partially inhibit Nitrobacter. A total nitrite-nitrogen concentration of 20 mg/1 would require a pH of 5.0 to shift the equilibrium to result in enough free nitrous acid to inhibit Nitrobacter. Verstraete et al. (1977) reported similar inhibitor concentrations of un-ionized ammonia and nitrite as Anthonisen et al. (1976). Since the pH of solution affects the equilibrium concentrations of ionized and un-ionized ammonia and nitrite, inhibition of nitrification is therefore a function of pH.

Other compounds in addition to free ammonia and nitrous acid have been reported to be inhibitory or toxic to nitrifiers (Lees, 1952; Aleem and Nason, 1960; Tomlinson et al., 1966; Loveless and Painter, 1968; Downing et al., 1964; Skinner and Walker, 1961; Hockenbury and Grady, 1977; Meiklejohn, 1954).

Lists of these compounds and metals that are inhibitory to nitrifiers are located in the above eight references. Hockenbury and Grady (1977) reported many organic compounds, especially nitrogenous organic compounds, that were inhibitory to nitrification. However, different compounds inhibited nitrification at different concentrations. Tomlinson and Snaddon (1966), stated that nitrifiers are less sensitive to heavy metals in activated sludge suspensions than in pure cultures although the same generalization could not be made for organic compounds. A possible reason for this phenomenon is that biological flocs can bioadsorb heavy metals but not organic compounds. Downing et al. (1954) and Mulbarger (1971) observed the acclimation by nitrifying bacteria to some organic inhibitors. Heukelekian (1942 and 1947) investigated the effects of carbonaceous materials on nitrification in activated sludge and trickling filter processes. It was reported that nitrification was not significantly affected by carbonaceous material with domestic wastewater except when the heterotrophic bacteria reduced the dissolved oxygen to a limiting level (e.g., 1.0 mg/1). However, Weng and Molof (1974) reported that for a RBC, a COD concentration greater than 50 mg/1 completely inhibited nitrification. The inhibition reported by Weng and

Molof (1974) was most probably due to a very low oxygen concentration (DO$\sim$0) and not a high COD concentration.

Process Parameters

The performance of RBC processes has been investigated by many researchers to determine what parameters affect the process. The predominant parameters and conditions that have been examined are:

(1) Temperature
(2) pH
(3) Rotational Speed
(4) Dissolved Oxygen
(5) Biofilm
(6) Media Surface Area
(7) Hydraulic Loading
(8) Influent Substrate Concentration
(9) Organic Loading
(10) Type of Waste
(11) Number of Stages
(12) Effluent or Sludge Recycle

Some of the parameters that have been investigated are interdependent and will be considered together.

Temperature

Temperature is a parameter of fundamental importance to the growth of micro-organisms. The metabolic rate of micro-organisms, diffusivities, mass transfer coefficients and the saturation concentration of oxygen are all dependent on temperature. Ellis and Bananga (1976) reported an increase in temperature from 11 to 27°C which increased BOD_5 removal from 90 to 94%; the average influent BOD_5 was 240 mg/l and the HRT was 3 hrs. Antonie (1976) indicated that temperature increases above 13°C did not improve organic removal; this conclusion was based on 12 different RBC municipal wastewater treatment plants. Benjes (1977) showed results from several different studies that indicated no improvement occurred in BOD_5 removal above 13°C and in ammonia oxidation above 16°C. Lue-Hing et al. (1976) reported that an increase in temperature from 10°C to 20°C increased the oxidation in ammonia from 15.6 to 43.5 lbs ammonia-nitrogen/d/1000 ft^3 of media volume; the average influent of ammonia-nitrogen was 780 mg/l. The results reported by Lue-Hing et al. (1976) may be compatible with those reported by Benjes (1977). In summary, the removal of BOD_5 and the oxidation of ammonia were independent of temperature when the temperature was above 13°C and 16°C, respectively.

pH

No experimental results have been found by the author where pH of the mixed liquor in a RBC process was the only variable tested. Several researchers have, however, observed increases in pH across RBC stages (Birks and Hynek, 1971: Hao and Hendricks, 1975A; Hudson et al., 1976). Hao and Hendricks (1975A) observed an increase in pH simultaneously with a decrease in alkalinity. Birks and Hynek (1971) reported an increase in pH from 6.8 to 7.8 across a four stage RBC system. Hudson et al. (1976) stated that the pH increased from 5.2 to 7.7 across a two stage RBC system. A possible reason for this phenomenon may be that carbon dioxide was being stripped from solution due to the rotating media. The agitation of the wastewater resulting from the rotation of the discs stripped carbon dioxide out of the wastewater and released it to the atmosphere. The stripping of the carbon dioxide results in a shift in the carbonate equilibrium. The reaction results in a production of hydroxide ions which increases the pH. The magnitude of the increase would therefore be a major function of the alkalinity of the wastewater and the removal rate of the carbon dioxide. The removal rate of the carbon dioxide is a function of the rotational speed

of the media, the media surface, the media configuration and the temperature. Temperature affects the diffusivity of carbon dioxide at the wastewater/air interface. The organisms oxidizing organic compounds produce carbon dioxide as a waste product. The net result of these conditions contribute to the determination of pH of a mixed liquor.

Rotational Speed

Rotation of the media, according to Antonie and Van Aacken (1971), also governs the intensity of contact between the bio-mass and the wastewater, and the rate of aeration. It was stated that a concentrated waste could be more effectively treated with higher rotational speeds. Antonie and Welch (1969) stated that increasing the rotation speed increases agitation and enables more substrate and dissolved oxygen to penetrate into the biomass. Antonie (1976) indicated that peripheral velocity (ft/min) may be a better measure of agitation than rotational speed (rpm).

Welch (1968) operated a pilot-scale RBC at rotational speeds of 5, 10, 20 and 30 rpm corresponding to peripheral velocities of 47, 94, 189 and 283 ft/min, respectively. With the same hydraulic loadings, the removal of organic material increased with each increase in rotational speed. Ellis and Banaga (1976) stated the optimum rotational speed for a bench-

scale RBC (disc diameter equal to 7.9 in.) was 4 rpm (8.3 ft/min). Chittenden and Wells (1971) operated a 3 stage pilot RBC system (disc diameter equal to 4 ft) at 3 rpm (38 ft/min) and obtained 49.7% BOD_5 removal. Holding all other variables at constant values, the rotational speed in the first stage was increased to 6 rpm (76 ft/min) and obtained a 64.5% BOD_5 reduction. Antonie (1976) stated that 60 ft/min was an approximate optimum peripheral velocity for a RBC treating a domestic wastewater. The varying results from these experiments indicate that other parameters and conditions affect the optimum rotational speed. The aeration capability of a rotating media may be the controlling condition that determines the optimum rotational speed when oxygen is limiting in a RBC process.

Dissolved Oxygen

If oxygen is limiting, the rate controlling step may be the oxygen diffusion rate between the air-liquid interface or the diffusion rate from the liquid into and through the biomass. The oxygen required by the biomass may be a function of the temperature, the type of waste or the dominate species of organisms. The RBC process will have a higher probability of being oxygen limited at high temperatures due to higher metabolic rates and the lower solubility of oxygen.

Bintanja et al. (1975) reported that with an RBC process higher concentrations of oxygen in the atmosphere above the discs resulted in higher COD reduction, lower sludge production and better settleability of the sludge. It was concluded that oxygen was limiting in the air system and organic substrate was limiting in the oxygen-rich system. Welch (1968) indicated that COD removal increased with dissolved oxygen concentration in the mixed liquor up to 1.5 mg/l and above 1.5 mg/l, the COD removal was constant. The COD removal also increased with rotational speed. The rotation of the media may have increased the agitation of the biomass and hence increased the oxygen available to the biomass. Either the rotation caused more turbulance which enabled more oxygen to penetrate the liquid-air interface or dissolved oxygen to penetrate further into the biomass. Torpey et al. (1972) observed higher organic removals with a RBC process, using an atmosphere rich in oxygen. Tomlinson and Snaddon (1966) concluded that a higher oxygen concentration above the biofilm would increase the diffusion of oxygen to the film base. Birks and Hynek (1971) stated that the thickness of a biofilm was regulated by the penetration of oxygen and organic substrate to the deepest layer of microorganisms.

Biofilm

Pretorius (1971) reported that biomass per unit area varied through a bench-scale RBC system. Nine single discs were operated in series; the mass of biomass per unit area increased from the first disc to the second and decreased on each of the seven remaining discs. Hudson et al. (1976), using shellfish processing wastewater, observed an increase in fixed biomass per unit area of media as the hydraulic retention time (HRT) was decreased. As the HRT decreased, the biomass per unit area reached a maximum at a HRT of 2 hrs and then the biomass per area became constant.

Media Surface Area

More biomass can obviously be supported on increased surface area which results in higher organic removals. Antonie and Welch (1969) expressed surface area in terms of surface area per unit total volume (the cylindrical volume enclosed by the outermost discs and disc diameter) the media occupied, called the specific curface area. The surface area is an important parameter and will be discussed in further detail in conjunction with organic loading.

Hydraulic Loading, Influent Substrate Concentration and Organic Loading

Many researchers have observed the effects of hydraulic

loading on the performance of a RBC. The removal of organic material was reported to increase with increased hydraulic detention time and then reach a constant level as hydraulic detention time continued to increase (Antonie, 1970; Chittenden and Wells, 1971; Ellis and Bananga, 1976; Hudson et al., 1976; Welch, 1968). Birks and Hynek (1971) and Welch (1968) each observed a decrease in the percent removal of COD as the influent COD concentration was increased. Ellis and Bananga (1976) concluded that influent organic concentration and hydraulic retention time are not independent in determining RBC performance and the use of organic loading (mass applied per media surface area per time) would incorporate the influent organic concentration, the hydraulic detention time and the media surface area. Ellis and Bananga (1976), Stover and Kincannon (1976), Torpey et al. (1971) and Welch (1968) observed increased organic loadings which resulted in decreased percent removals of organic substrate. Stover and Kincannon (1976) and Welch (1968) also reported that organic substrate removed (mass per time) increased with increased organic loading. As the organic loading continued to increase, the organic substrate removed reached a constant value. The organic loading at this constant value was considered to be the maximum capacity of the particular RBC

systems being studied. Welch (1968) increased the rotational speed and observed an increase in the maximum value for organic substrate removed. Steels (1974) reported effluent BOD_5 increased with increased organic loading. Torpey et al. (1971) used the organic loading concept in terms of ammonia. He observed similar correlations to that of organic loading.

Type of Waste

The type of wastewater treated has an impact on the performance of a RBC. Stover and Kincannon (1976) operated a RBC process with a synthetic sucrose wastewater and then with a slaughterhouse and meat processing wastewater. The percent COD removals were 96 and 75, respectively. The main difference between the wastewaters was the rate of biodegradability.

Number of Stages

Stover and Kincannon (1976) used a six stage RBC system for the experiments. The majority of the removal (i.e. 90%) for the synthetic sucrose substrate occurred in the first stage, with very little occurring in the last five stages. For the slaughterhouse wastewater, only 35% was removed in the first stage and then approximately 8% was removed in each of the following five stages. The substrate removal rates (based on COD) for the sucrose and slaughterhouse substrates

under the same conditions were 2.5 and 0.07 d^{-1}, respectively. A wastewater that has a high substrate removal rate can have a high loading rate.

The number of stages required, according to Famularo (1976), is a function of the wastewater concentration. Famularo (1976) stated that removal rate was a function of effluent organic concentration; therefore, low influent concentrations could be treated most efficiently by a series of RBC systems to simulate plug flow. In addition, at high organic concentrations removal rates were independent of substrate concentration until the concentration decreased enough to effect the removal rate. It was concluded that for high strength wastewater, one large first stage RBC should be used and then followed by a sequence of small stages approaching plug flow.

Chittenden and Wells (1971) concluded that the first stage of a three stage system (influent BOD_5 equal to 225 mg/1 mg/1) removed the majority of the BOD_5. The cumulative BOD_5 removal by the three stages were 80%, 83% and 83%, respectively. Torpey et al. (1971) operated a ten stage RBC system and observed removal of BOD_5 and COD by each stage; the influent BOD_5 concentration was 124 mg/1. The cumulative BOD_5 removals through the 10 stages were 34, 52, 65, 77, 85, 86, 89, 90, 93

and 93%, respectively. The major portion of the removal occurred in the first 5 stages. It was noted that ammonia oxidation did not occur until stage seven. These two cases demonstrate the concept that Famularo (1976) presented.

Effluent or Sludge Recycle

The recycle of sludge and effluent has been investigated as a possible means of improving performance of a RBC process. Lue-Hing et al. (1976) reported minimal increase in organic removal by recycling between 49 and 186% of the effluent flow. Ellis and Bananga (1976) reported similar results recycling 100% of the effluent flow. Welch (1968) recycled sludge from a clarifier following a RBC and observed an increase in performance. This increase may have been due to the 1500 mg/l increase in the mixed liquor volatile suspended solids in the RBC.

The shearing forces that occur in a RBC system contribute to the removal of the biofilm from the media. The organisms that remain attached multiply and produce a new biofilm which again sloughs (Hudson et al., 1976; Welch, 1968; Bintanja et al., 1975; Ellis and Bananga, 1976). This cycling sets up an average cell residence time in a RBC process. Cell residence time in a RBC process, to the author's knowledge, has not been investigated. In this study, a bench-scale RBC

system was set up to evaluate cell residence time as well as organic removal and nitrification.

CHAPTER III

EQUIPMENTAL SYSTEM AND PROCEDURES

Experimental System

Reactor Description

The RBC system used in this investigation is shown in a photograph in Figure 3. Detailed schematic diagrams of the RBC reactor and the total system are presented in Figures 4 and 5, respectively. The reactor was constructed of a clear acrylic cylinder divided along the longitudinal axis with one-half comprising the main unit and the other half used as a cover. The acrylic cylinder was 40.6 cm long, 30.5 cm in diameter and had a 1.27 cm wall thickness. The single stage reactor had a 5.34 l liquid capacity at a maximum liquid depth of 8.9 cm. The unit contained 12 discs which were 0.32 cm thick and 25.4 cm in diameter. The discs were set at cm spacings on a stainless steel shaft with a maximum submergence depth of 7.6 cm. The total wetted disc surface area provided for biological growth was equal to 1.0 m^2 with a hydraulic loading of approximately 70 $l/d \cdot m^2$ and a liquid retention time of approximately 2.0 hr. A 1.27 cm clearance was provided between disc edges and the contoured basin bottom to avoid

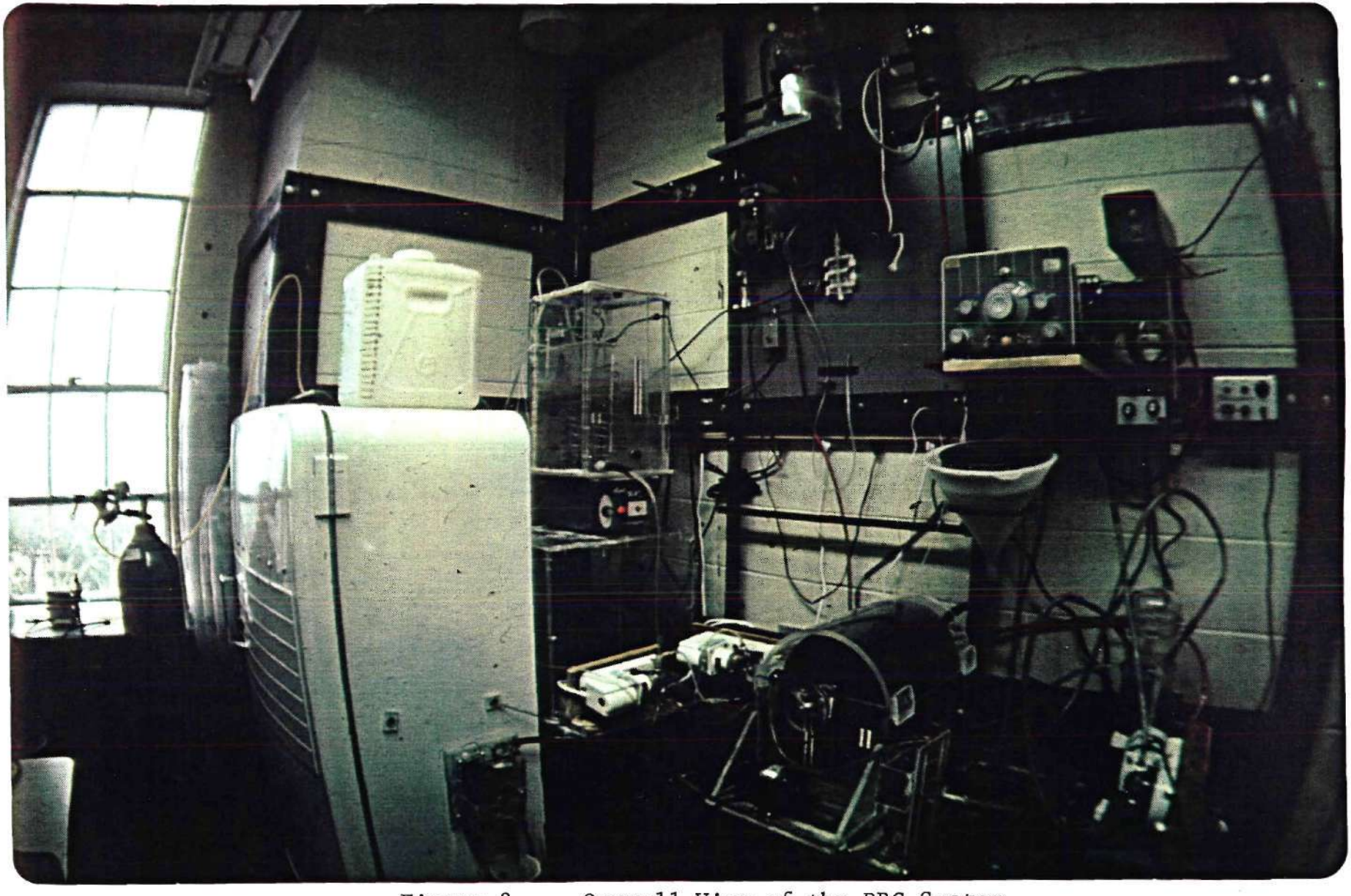

Figure 3. Overall View of the RBC System

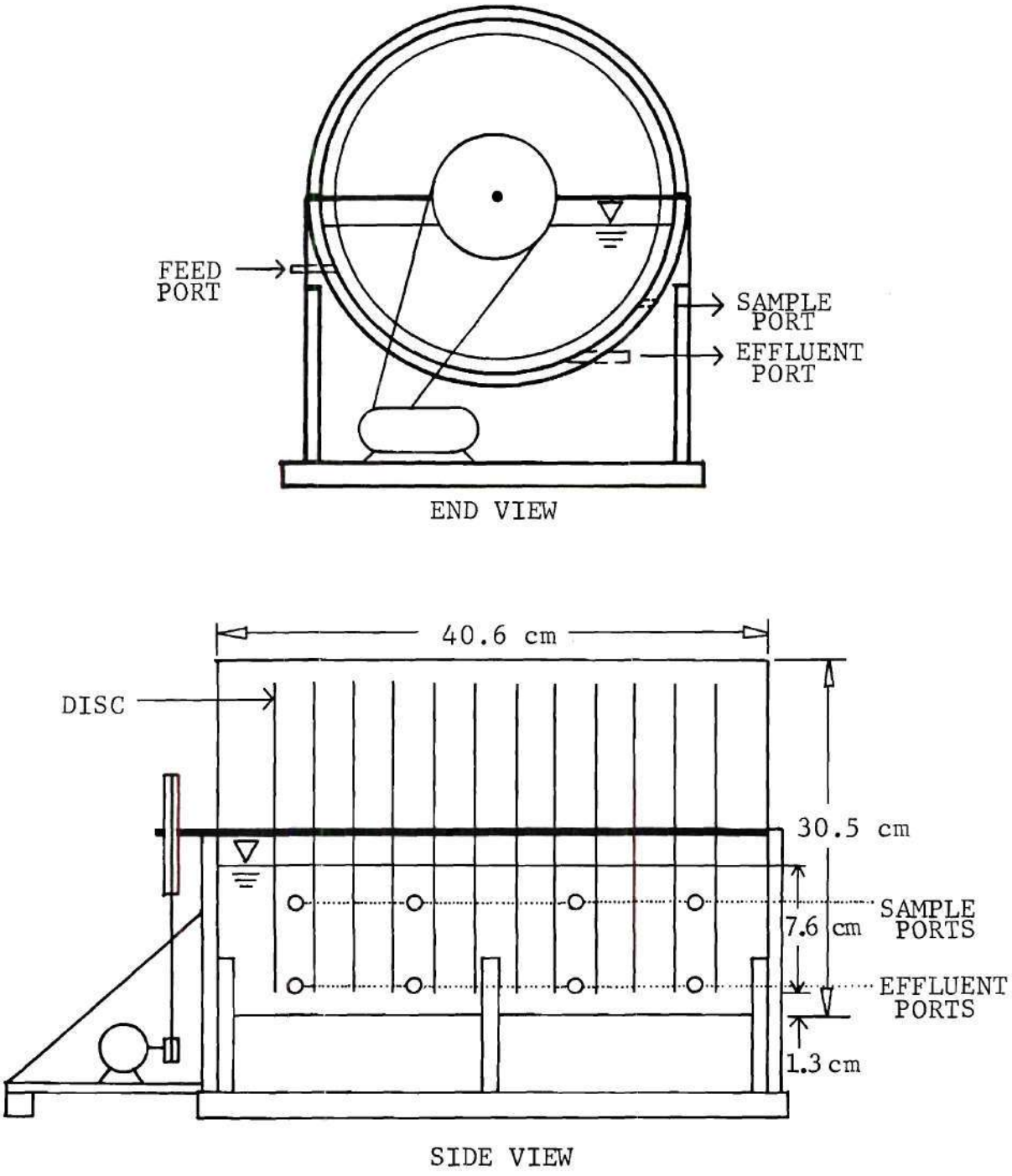

Figure 4. Schematic Diagram of the Laboratory Scale Rotating Biological Contactor

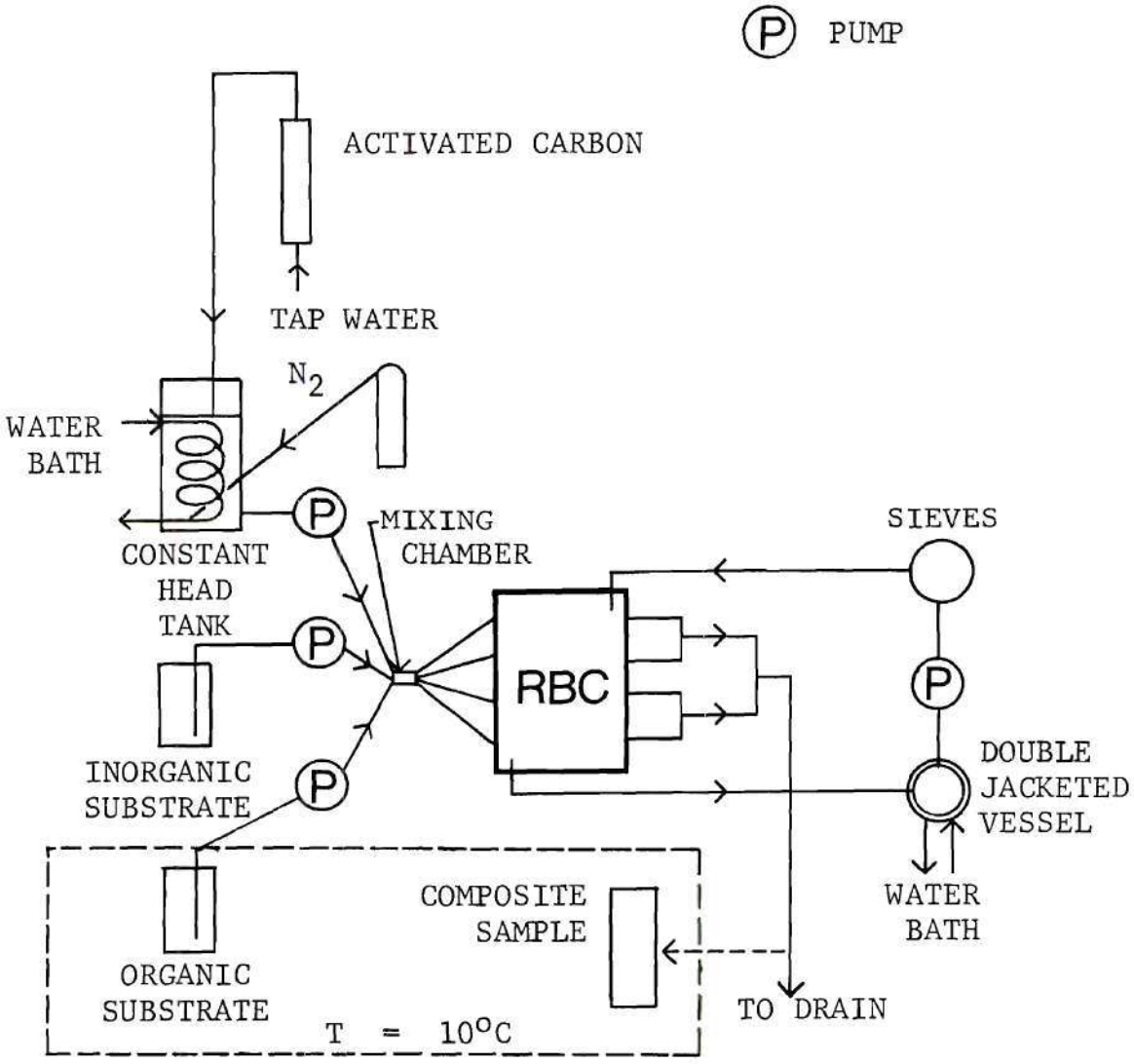

Figure 5. Schematic Diagram of the RBC System

short circuiting. The acrylic cover was used to minimize evaporation losses and to eliminate drastic changes in temperature. A variable speed electric motor, Type NSH-12R Bodine, with a capacity of 1/50 hp was mounted on the side of the disc frame assembly and connected with an "O" ring to the stainless steel shaft.

As shown on the schematic diagram in Figure 5, tap water was dechlorinated in an activated carbon column and held in a constant head tank for use as dilution water. Nitrogen gas was continuously bubbled through the dilution water in the constant head tank with a diffuser stone to strip dissolved oxygen from the water. The temperature was controlled in the tank with a submerged coil of stainless steel tubing through which water was circulated from a temperature-controlled water bath. Organic substrate was stored in a 20 l carboy kept at $10^{o}C$ in a refrigerator while the inorganic substrate was stored at room temperature in a 20 l carboy. The dilution water and the organic and inorganic substrates were pumped to a mixing chamber with variable speed FMI, model RRP, lab pumps. The three streams were mixed in the mixing chamber which consisted of a 7.6 cm long piece of plastic tubing with a 2.5 cm diameter and two rubber stops. The effluent of the mixing chamber

entered the RBC reactor through four feed ports shown in Figure 4.

To assure the reactor mixed liquor was well mixed, a circulation system was installed (see Figure 6). The mixed liquor was pumped from one end of the reactor through a 0.4 l double jacketed vessel and discharged into a sieve assembly. Water from the temperature controlled water bath was circulated through the outer jacket of the double jacketed vessel. The sieve assembly consisted of a No. 20 sieve stacked on top of a No. 80 sieve with both sieves mounted in a large funnel. The filtered wastewater flowed by gravity into the opposite end of the reactor. The circulation flow was 1.0 l/min; therefore, the mixed liquor was recycled every six min. With a hydraulic retention time of 120 min (the total system volume including the circulation system was 6 l) the mixed liquor was theoretically circulated through the circulation system 20 times during every hydraulic retention time.

The reactor effluent was discharged through four ports, as shown in Figure 4. Effluent composite samples were collected daily at 15 min intervals with an automatic sampler. The composite sample reservoir was located in a refrigerator at $10^{o}C$.

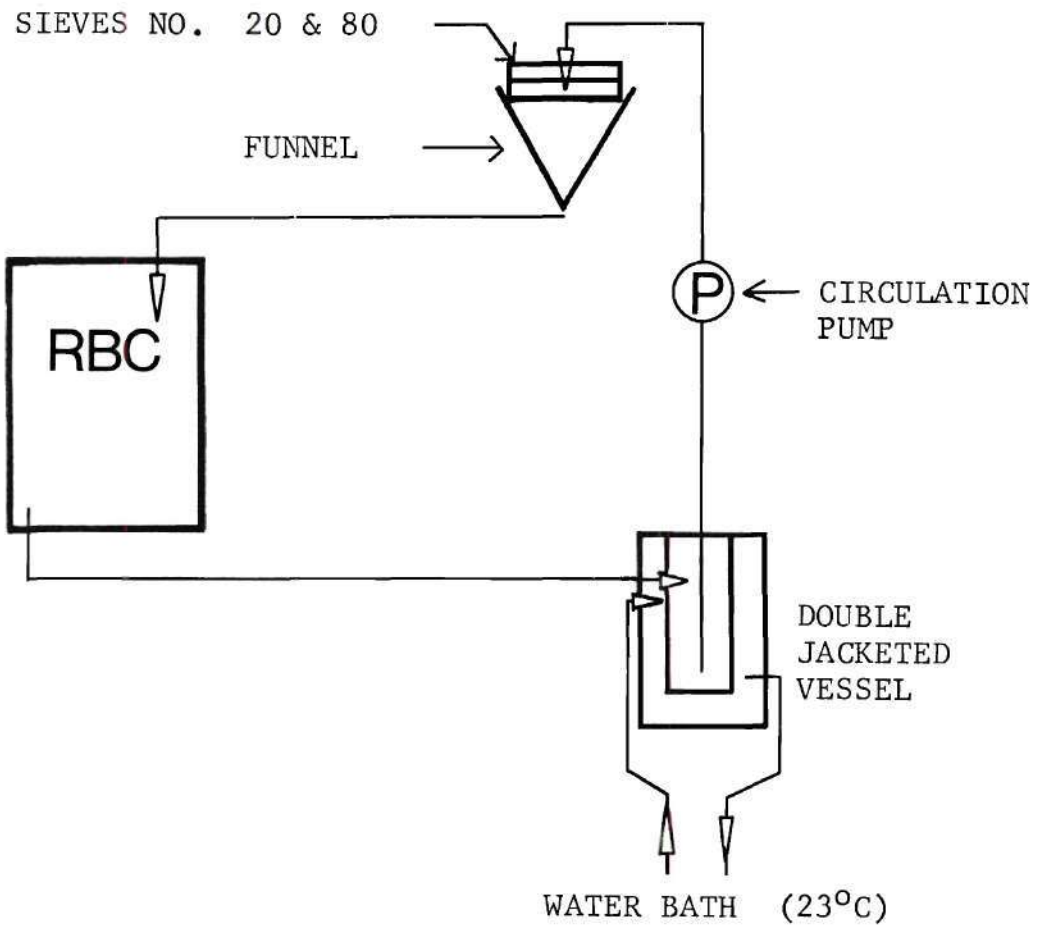

Figure 6. Circulation System

Substrate

Since it was desired to study the applicability of the RBC process for nitrification of a secondary effluent, a synthetic wastewater was prepared to simulate the soluble portion of a typical secondary effluent. A synthetic wastewater eliminated any variations in the wastewater strength that otherwise would exist if an actual secondary effluent was used. The use of a synthetic substrate also enabled a more effective evaluation of system performance since the nitrogeneous species of interest could be monitored accurately.

From the data presented by Mueller et al. (1958), Painter et al. (1961), Painter (1971) and Rebhum and Manka (1971), representative organic and inorganic compounds were chosen to simulate the soluble fraction of a typical secondary effluent. These compounds are presented in Tables 3 and 4 with the concentrations utilized for this study. As listed in Table 3, ferric chloride was included in the organic substrate to avoid precipitation of the iron in the inorganic reservoir. To assure the consistency of the synthetic wastewater and to minimize the pumping of large volumes, the wastewater was made in concentrated batches. The concentrated organic substrate was placed in plastic bottles in 200 ml portions and frozen until needed. The 200 ml concentrated

Table 3

Organic Substrate Components

Compounds	Concentration (mg/l)
Arabinose	4.7
Galactose	4.7
Sucrose	11.9
Xylose	4.7
Formic Acid (1.22 g/ml)	1.46
Acetic Acid (1.05 g/ml)	0.76
Propionic Acid (0.99 g/ml)	0.83
Butyric Acid (0.96 g/ml)	0.80
Valeric Acid (0.94 g/ml)	0.76
Lactic Acid (1.206 g/ml)	1.45
Citric Acid	2.10
Benzoic Acid	0.63
Instant tea (Nestea*)	10.4
Phenol (1.07 g/ml)	0.11
COD	45.3
$FeCl_3.6H_2O$	0.45

*Trade Mark of Nestle Co., Inc., New York

Table 4

Inorganic Substrate Components

Compounds	Concentration (mg/1)
$MnSO_4.H_2O$	0.56
$ZnCl_2$	0.95
$CuSO_4.5H_2O$	0.3
$CoCl_2.6H_2O$	0.38
$Na_2B_4O_7.10H_2O$	8.8
KH_2PO_4	8.5
K_2HPO_4	21.75
$Na_2HPO_4.7H_2O$	33.4
$(NH_4)_2SO_4$	92.2
NH_4Cl	1.7
$CaCl_2$	5.0
$MgSO_4$	5.0
$NaHCO_3$	168.0
Na_2CO_3	212.0
Na_2SO_4	200.0

organic substrate (COD = 45,000 mg/1) was diluted to 15 1 with distilled water and pumped into the mixing chamber at 5 ml/min. The total flow to the reactor of 50 ml diluted the organic substrate to a COD equal to 60 mg/1; the dilution water and inorganic substrate flow rates were 40 ml/min and 5 ml/min, respectively. When higher influent COD concentrations were required, the concentration of the organic substrate was increased.

The nitrogen concentration of the inorganic substrate was 2000 mg/1 as ammonia-nitrogen. Two of the concentrated inorganic substrate were diluted to 20 1 and pumped into the mixing chamber at approximately 5 ml/min. The total flow of approximately 50 ml/min diluted the influent nitrogen to a concentration of approximately 20 mg/1 as ammonia-nitrogen.

The influent COD to nitrogen ratio ranged from 3.03 to 14.3. The influent COD to phosphorous ratio ranged from 6.25 to 33.3. Antonie and Welch (1969) stated that an acceptable COD to nitrogen and COD to phosphorous ratios were 30.3 and 143, respectively. According to these accepted ratios, nitrogen and phosphorous were not limiting substrates for the heterotrophic populations in this study.

Operational Parameters

The RBC system used in this study was a single stage

unit with no sludge recycle. The media surface area was constant. The temperature and pH of the mixed liquor, the rotational disc speed and the hydraulic retention time were controlled at constant values.

Temperature of the mixed liquor was controlled at 23°C to 24°C with maximum fluctuations of 2°C. Water from a temperature controlled water bath was pumped through the stainless steel coiled tubing in the constant head tank and the outer jacket of the double jacketed vessel in the circulation system to control the temperature of the mixed liquor. A magnetic stirrer was used in the constant head tank to assure good temperature distribution of the dilution water. The cover used for the reactor reduced evaporational losses and therefore helped to control the temperature. The pH in the mixed liquor was controlled by adjusting the pH in the organic and inorganic substrates to achieve a mixed liquor pH of 7.2. Since reactor pH was controlled by adjustment of the pH of the influent substrate, the buffer capacity of the substrate was very low. The alkalinity supplied in the substrate was equal to 27.4 mg/l as $CaCO_3$. The rotational speed of the discs was held constant at 30 rpm and the peripheral velocity of the discs was 23.9 m/min. The hydraulic retention time was held at approximately 120 min by controlling the

influent flow at 50 ml/min. Since the disc surface area was also constant, the hydraulic loading was approximately 70 $l/d \cdot m^2$ throughout the study.

The dissolved oxygen concentrations in the mixed liquor and in the film were not controlled. The oxygen transfer coefficient, $K_L a$, for the RBC reactor with no biomass on the discs and operating at 30 rpm was determined by McIndoe (1976) to equal 0.125 min^{-1} at a temperature of 20° C. The dissolved oxygen concentration in the dilution water was controlled by bubbling nitrogen gas in the constant head tank with a stone air diffuser to strip oxygen from the water. The dilution water was controlled at a dissolved oxygen concentration of approximately 2 mg/l to simulate the dissolved oxygen concentration of a typical secondary effluent. A YSI Model 54 oxygen meter was used daily to monitor the dissolved oxygen in the dilution water and the RBC mixed liquor.

The influent concentration of soluble nitrogen species was constant throughout the study at a concentration of approximately 20 mg/l. Since the hydraulic loading rate was a also constant, the nitrogen loading rate was approximately 1.4 $gm/d \cdot m^2$. The influent COD concentration was changed during the study with the range of 57 mg/l to 193 mg/l. Therefore, the organic loading varied between 3.9 $gm/d \cdot m^2$ and 13.7 $gm/d \cdot m^2$.

The synthetic wastewater composition was held constant for the duration of the study with one exception. The source of nitrogen normally used was ammonium ions, except in the second experiment where glycine was the sole nitrogen source. The concentration was adjusted to an influent nitrogen concentration of 20 mg organic-nitrogen/l.

The depth of the biofilm on the discs was controlled only by the scraping procedure used to control the cell residence time, which will be discussed in detail in following sections.

Experimental Procedures

Sampling Procedures and Data Collection

Influent substrate samples were withdrawn from the respective substrate reservoirs and analyzed for COD and ammonia-nitrogen. In one experiment, the organic substrate was also analyzed for organic nitrogen since the nitrogen source (i.e. glycine) was added to this substrate. In addition, 24 hr effluent composite samples were analyzed for COD, ammonia-nitrogen, nitrate-nitrogen, TSS and VSS. In the experiment with glycine as the nitrogen source, the effluent was analyzed for organic nitrogen and nitrite-nitrogen.

Effluent suspended solid concentrations and three addi-

tional suspended solid fractions were collected and analyzed. The biomass on the discs was removed at periodic intervals; the scraping intervals will be discussed later. This fraction was referred to as SS_D. Another fraction of suspended solids was collected on the No. 20 and 80 sieves in the circulation system and this fraction was referred to as SS_C. When the discs were scraped, the suspended solids collected on the sieves during the one to three d period between scrapings were washed into a plastic storage bottle. Another fraction was the suspended solids in the mixed liquor that had not been removed from the RBC or retained on the sieves in the circulation system. During the period of disc scraping, the mixed liquor was pumped into a 10 l carboy and then poured back into the reactor through a No. 100 sieve. The solids retained on the sieve were collected and analyzed; this fraction was referred to as SS_M. All four fractions were each analyzed for TSS and VSS. To analyze for organic-nitrogen, 20 percent of each of the four fractions were mixed prior to analysis. Therefore, organic-nitrogen of the solids was determined on a composite of the solids wasted from the system.

With the COD and the nitrogen data, the removal efficiencies of the system in terms of organic carbon and nitrogen removal were calculated. The nitrogen species data were also

used to make a nitrogen balance and account for the total influent nitrogen in the effluent fractions.

The influent organic and inorganic substrate flow rates were determined by weighing the carboys once a day, converting the mass value to a volumetric value and calculating the change in volume over a period of time. The dilution water and wastewater effluent flow rates were obtained by measuring the flow into a graduated cylinder over a period of time. Disc rotational speed was monitored by marking a line on the wheel outside the disc assembly shaft and counting the number of revolutions per minute. Temperature was monitored with thermometers in the reactor and the constant head tank.

The total mass of biofilm was determined at the end of each of the first three experiments. This was accomplished by scraping a representative number of disc sides in each set, either a whole side or one quarter of a side. The age distribution of the biomass due to the scraping procedure was taken into consideration. These data were used to determine the mean cell residence time.

Control of the Cell Residence Time

The cell residence time of the RBC process was controlled by scraping portions of biofilm from the system. The actual cell residence time was not known until the end of the

experiment. Therefore, the term operational cell residence time (θ_{op}) was developed and used in process operation and control during the studies. The θ_{op} parameter was, however, developed more for communication purposes than for actual definition of the process.

The procedure used to control the cell residence time is described below. The twelve discs of the RBC reactor were divided into four sets with three discs in each set. Each side of a disc in each set was assigned a number from one to six. A set with the numbered sides is presented in Figure 7. The disc sides with the same number were evenly distributed along the longitudinal axis of the reactor and provided for an even distribution of the biomass in the reactor.

During the first experiment, the same numbered sides were scraped every third day, i.e. four sides were scraped every three days. A complete cycle was made in 18 d; and therefore, the θ_{op} for this experiment was assigned the value of 18 d. A θ_{op} of 12 d required that four identical numbered sides be scraped every other day. Therefore, a cycle would be completed in 12 d.

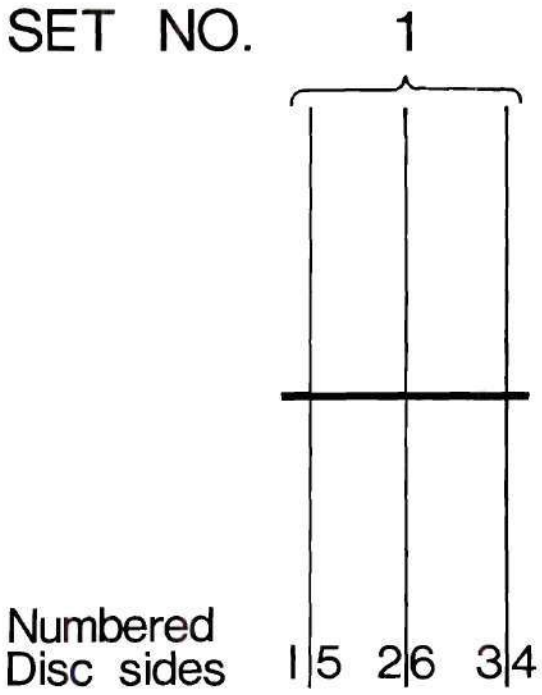

Figure 7. Numbering System for Disc Sides

Analytical Techniques

Chemical Oxygen Demand

The concentration of organic matter in the wastewater samples was measured by the Chemical Oxygen Demand (COD) test as described in Section 508 of Standard Methods (1977). The alternate procedure requiring the use of 0.025 N potassium dichromate and 0.010 N ferrous ammonium sulfate was used. Nitrite interference was eliminated by adding sulfamic acid to the standard dichromate solution. All unknown, blank and standardization samples were filtered and analyzed in duplicate. The precision was evaluated with standard solutions of potassium acid phthalate using four replicate samples and is shown in Figure 8. The coefficients of variation were within acceptable limits.

Ammonia Nitrogen

Ammonia analyses were performed using a Orion model 95-10 ammonia specific ion electrode together with a Leeds & Northrup pH/mv meter and a Sargent (Model SRG) strip chart recorder. A standard addition procedure was used to determine all ammonia-nitrogen concentrations and each sample was run in duplicate. The ammonia electrode responded to ammonia in accordance with the Nernst equation. A final equation for ammonia nitrogen is presented below and expresses the unknown

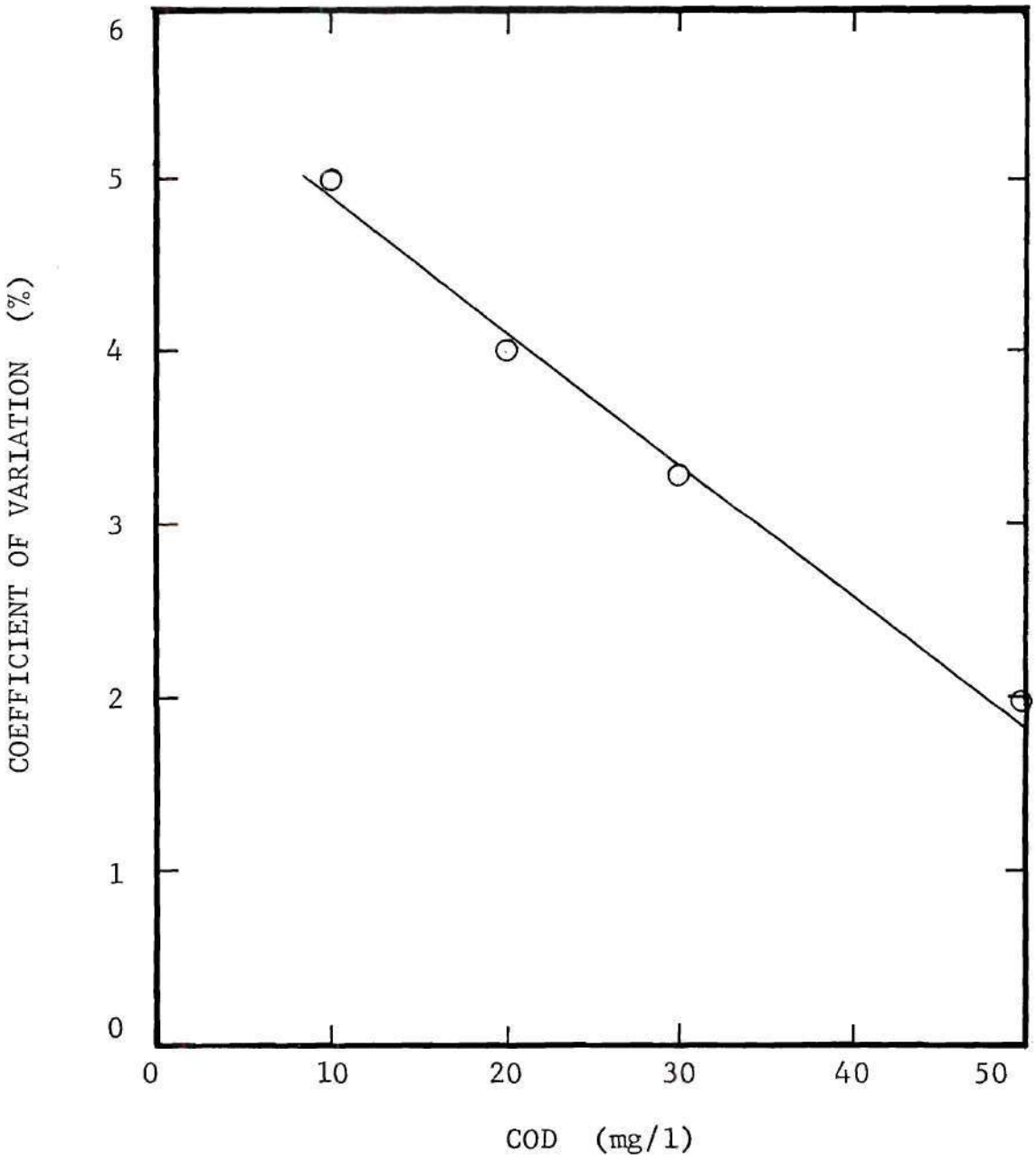

Figure 8. Precision of COD Test

ammonia-nitrogen concentration, N_1, in terms of the standard ammonia-nitrogen concentration, N_{std}; the difference in the electrode potential between the standard and unknown sample, E; and the electrode slope, S, a constant with units of mv.

$$N_1 = \frac{0.1\ N_{std}}{(10^{\frac{E}{S}}) - 1} \qquad (9)$$

Orion (1975) provided a table, based on a S equal to 59 mv at 25°C, to determine N_1. Standard ammonia-nitrogen concentrations were plotted against concentrations measured by using the probe and Orion (1975) table in Figure 9. The measured concentrations were higher than the actual concentrations. Instead of determining a new S value and using Equation 9, the slope of the line in Figure 9 was used. The measured concentration was divided by the slope, i. e. 1.25, to determine the actual ammonia-nitrogen concentration.

Nitrite Nitrogen

The diazotization method (Section 420 - Standard Methods, 1977) was performed on filtered composite effluent samples to determine nitrite-nitrogen. A minimum of two standard solutions of sodium nitrite were included with each set of samples. A typical standard curve determined with sodium nitrite is presented in Figure 10.

Nitrate Nitrogen

The chromotropic acid method (Section 419E - Standard Methods, 1977) was used for nitrate-nitrogen analysis. Three standard solutions of potassium nitrate were included with each set of samples. A typical standard curve determined with potassium nitrate solutions is shown in Figure 11.

Total Kjeldahl Nitrogen

The Technicon Corporation Industrial Method 28-69A was used to determine total Kjeldahl nitrogen concentration (TKN) of the suspended solids and the influent and effluent wastewater samples. The samples were digested in a continuous digester with an approximate detention time of 25 min. The released ammonia was measured by utilizing the berthelot reaction in which the formation of a blue-indophenol complex occurred when ammonia was reacted with sodium phenate followed by the addition of sodium hypochlorite (Ferrari, 1960). The organic nitrogen content in the solids was determined by subtracting the ammonia-nitrogen concentration from the TKN concentration. Equal portions of each of the solids fractions removed from the RBC system, excluding the solids in the effluent, were mixed together for organic nitrogen analysis. The mixture was blended prior to analysis to provide an average nitrogen content of the suspended

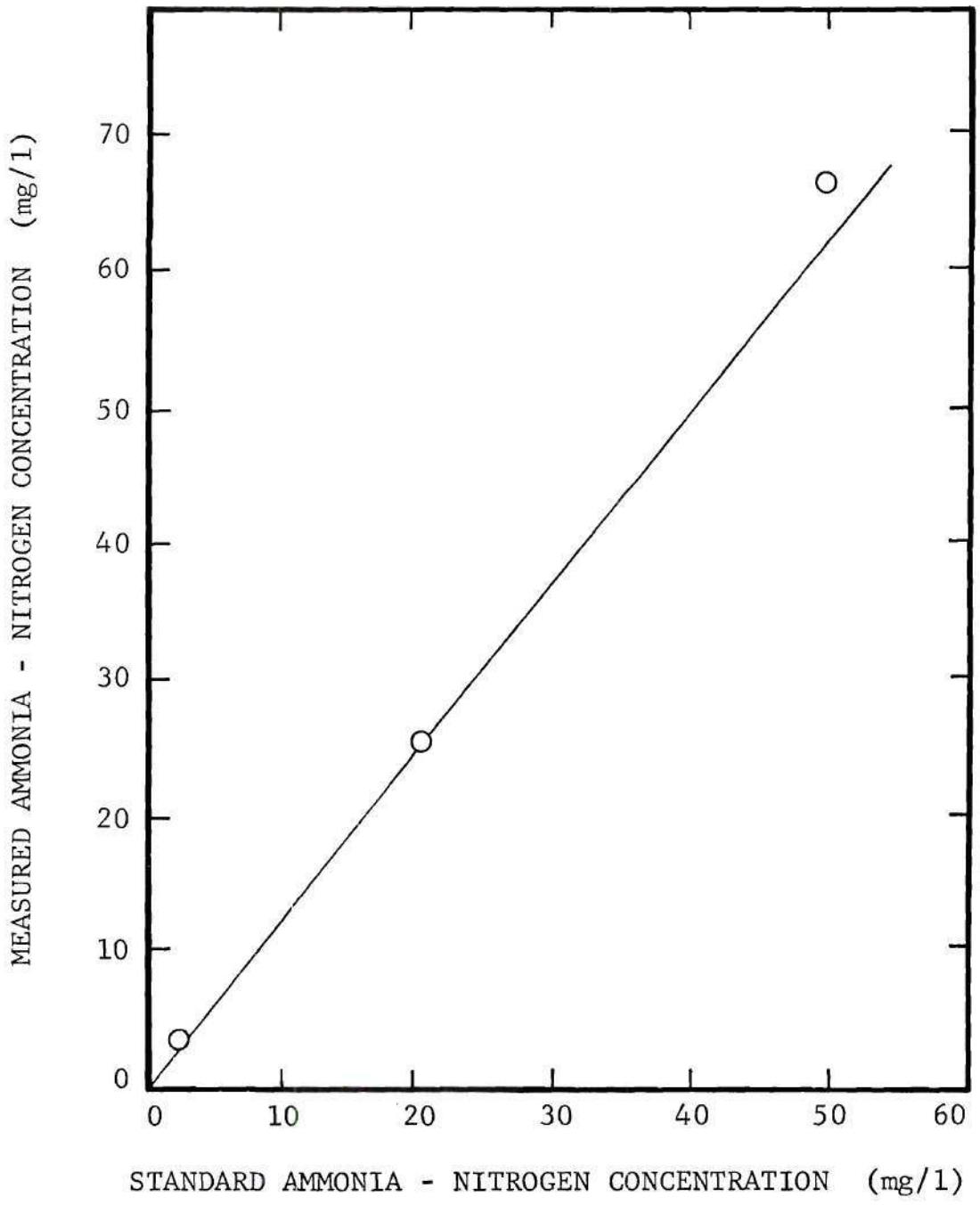

Figure 9. Response of the Ammonia Probe

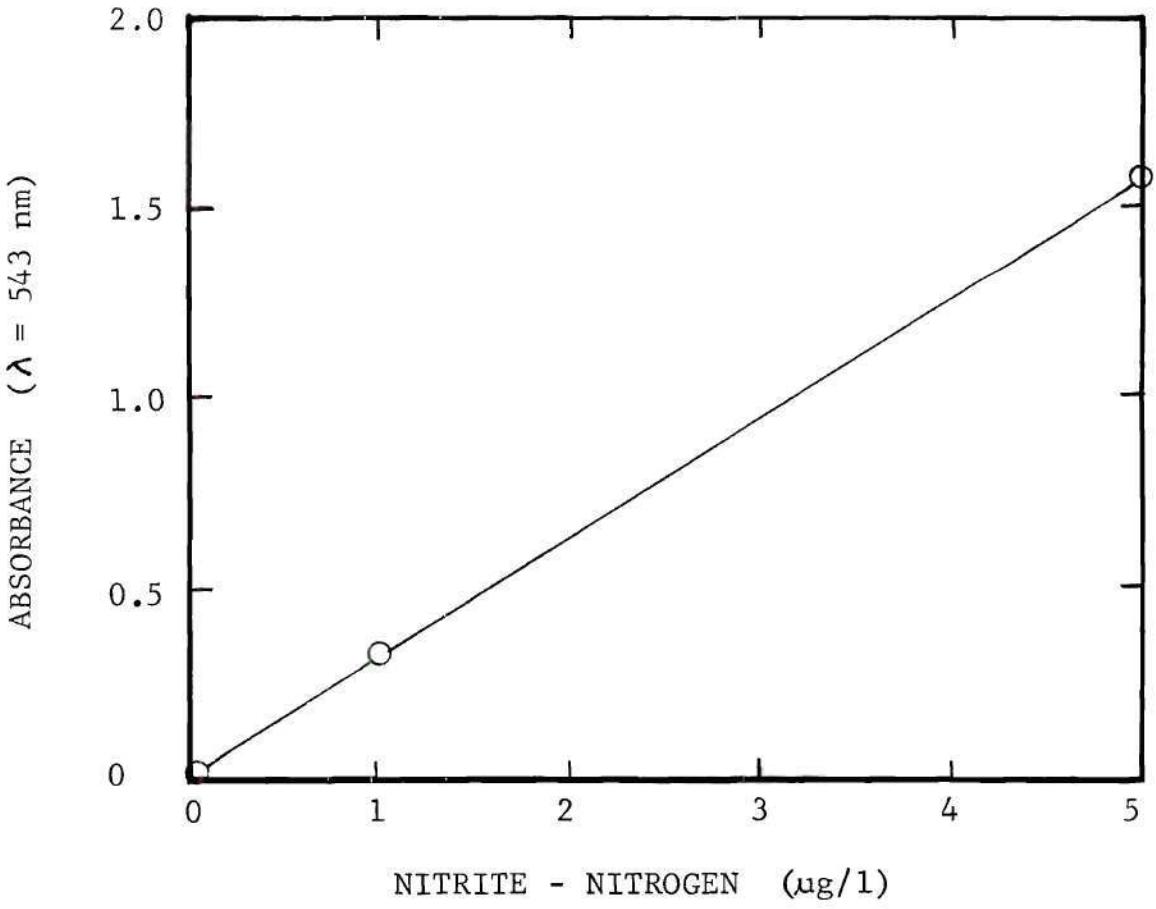

Figure 10. Typical Standard Curve for the Nitrite - Nitrogen Test

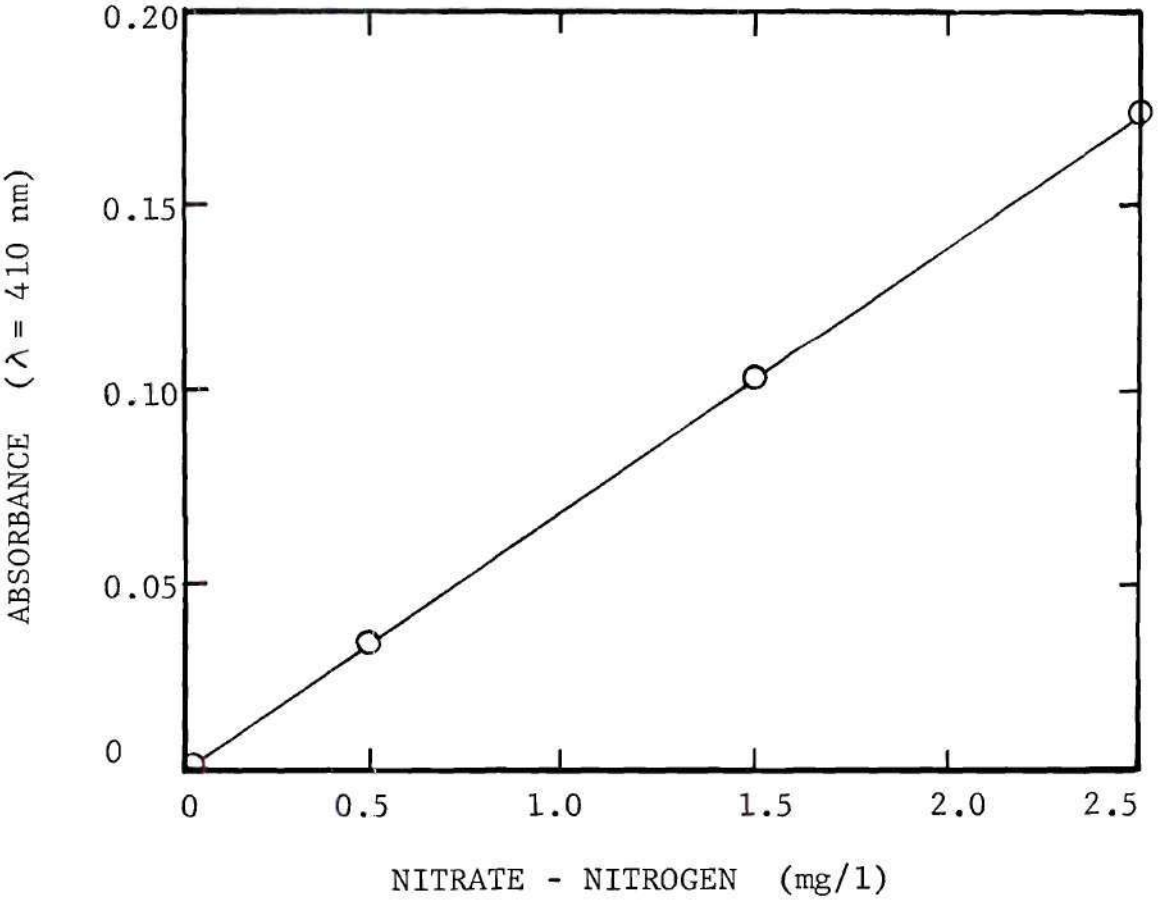

Figure 11. Typical Standard Curve for the Chromotropic Nitrate - Nitrogen Test

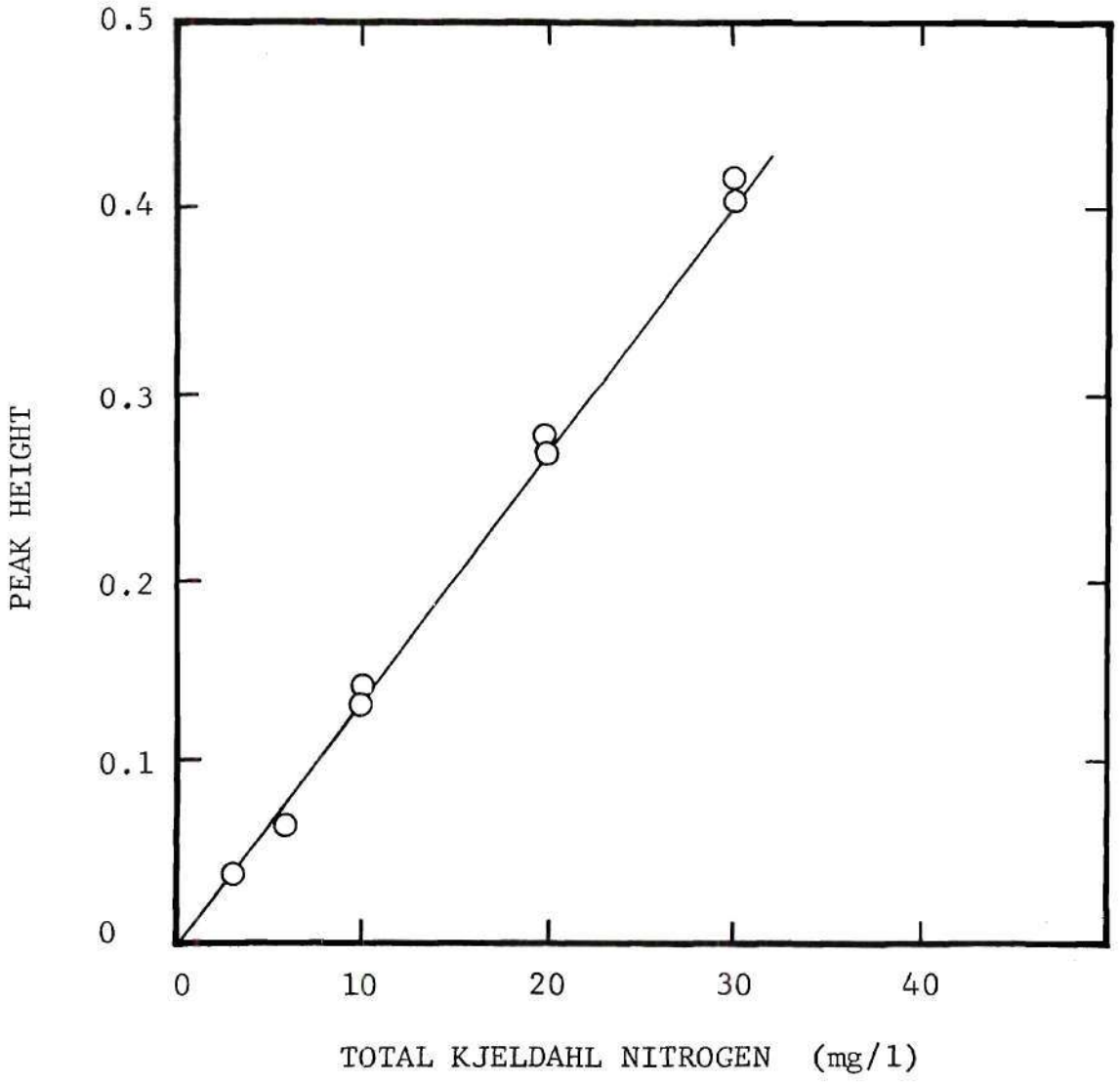

Figure 12. Typical Standard Curve for the TKN Test

solids. Glycine was used as a standard for the TKN analysis. A typical standard curve for the analysis performed with the Technicon Autoanalyser is presented in Figure 12.

Suspended Solids

The suspended solids concentrations were determined with Gooch crucibles and glass fiber filters as described in Section 208D of Standard Methods (1977). TSS and VSS analyses were performed on the effluent, the solids scraped from the discs, the solids retained on the No. 80 and 20 sieves, and the solids collected from the mixed liquor. To collect the solids from the mixed liquor, the liquid contents of the RBC reactor were poured through a No. 100 sieve. For the effluent, 250 ml of sample was filtered. The other three solid fractions were each brought to a volume of 200 ml and then 10 ml samples were filtered. All samples were analyzed in duplicate.

Description of Experiments

The objective of the research was to determine the capacity of an RBC system to oxidize organic matter and ammonia in a secondary wastewater and to determine the effects of θ_c on nitrification. The operation of the RBC was therefore divided into five phases.

Phase I was run with influent COD and ammonia-nitrogen concentrations of 62 mg/l and 15.6 mg/l, respectively; these concentrations are typical of secondary effluents from domestic wastewater treatment plants (Culp, Wesner and Culp, 1978). Normally, the θ_c in a RBC system nitrifying a secondary effluent is $\geq$30 d.(Antonie, 1976). To evaluate the RBC system at a lower and more critical θ_c, a θ_{op} value of 18 d was used.

For Phase II, the ammonia in the influent was replaced with an organic compound containing nitrogen, i.e. glycine, at a concentration of 23.1 mg-nitrogen/l. The nitrogen source was changed to determine the effect of the hydrolysis of organically-bound nitrogen on nitrification in the RBC process. The influent COD during this phase was 84 mg/l and the θ_{op} was held at 18 d. In subsequent phases, ammonia was the only influent source of nitrogen. During Phase III, the influent ammonia-nitrogen was at a concentration of 17.5 mg/l. The influent COD was increased to 90 mg/l to determine the effect of a higher influent COD on nitrification and COD removal. The θ_{op} was decreased to 12 d to evaluate these effects at a lower and more critical θ_c value.

During Phase IV, the influent COD concentration was increased to 136 mg/l to observe the effect of increased organic

matter concentration on nitrification. The θ_{op} and influent ammonia-nitrogen concentration remained at 12 d and 20.5 mg/l, respectively.

Phase V was subdivided into two parts, i.e. parts A and B. The θ_{op} and influent ammonia-nitrogen concentration were 12 d and 20.2 mg/l, respectively. The influent COD concentration during part A was increased from 136 mg/l to approximately 172 mg/l.

During part B of phase V, the influent COD concentration was increased by approximately 8 mg/l·d for 8 d and then held at approximately 200 mg/l for the remainder of the study, i.e. 3 d. The continual increase in COD was to simulate a failing secondary treatment system and to observe what effect the increased COD would have on nitrification. The θ_{op} and influent ammonia-nitrogen concentrations were 12 d and 18.0 mg/l, respectively.

The overall objectives of the above were to determine the effects of organic matter concentration, nitrogen source and cell residence time on nitrification. A summary table of the operating conditions is presented in Table 5. The data collected by Cruz (1977) were evaluated as well as the data from the present study; therefore, the operating and data conditions used by Cruz (1977) are included in Table 5.

Table 5

Summary Table of RBC Operating Conditions

	Influent Nitrogen		Influent COD	θ_{op}
Phase	Nitrogen Source	Concentration (mg-nitrogen/l)	Concentration (mg/l)	(days)
I	Ammonia	15.6	62	18
II	Glycine	23.1	84	18
III	Ammonia	17.5	90	12
IV	Ammonia	20.5	136	12
V A	Ammonia	20.2	172	12
V B	Ammonia	18.0	190	12
Run[1]				
1	Ammonia	18.8	57	6
2	Ammonia	19.8	63	12
3	Ammonia	19.6	60	—[2]

Legend
1) Data from Cruz, 1977.
2) No scraping performed during Run 3.

CHAPTER IV

RESULTS AND DISCUSSION

Operating Conditions

The conditions under which the RBC was operated during the five phases are listed in Table 6. The data collected by Cruz (1977) using the same RBC system are also included. The mean, standard deviation and number of data points are given for each condition. The mean temperature of the RBC mixed liquor during each experiment did not vary significantly. The arithmetic mean of the pH values varied minimally, with only two exceptions. During run 2 (Cruz, 1977), the mean pH was 6.8; however, according to Haug and McCarty (1971) nitrifiers can acclimate to this pH. The other exception was a 2 hr pH shock during run 1 (Cruz, 1977); this will be discussed later. The small fluctuations in the mean temperatures and pH values were considered to be negligible; therefore, the temperature and pH were considered to be constant during the experiments. The mixed liquor pH was approximately 1 unit higher than the influent pH. An increase in pH was also observed by Birks and Hynek (1971), Hudson et al. (1976) and Mulbarger (1971). This phenomenon is assumed to

Table 6

Operating Conditions During Phases I-V and Runs 1-3

PARAMETER	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
Temperature (°C)	23.3	24.9	24.4	23.8	22.7	21.3	21.4	24.9	23.5
	(1.2)	(1.5)	(0.8)	(1.7)	(0.9)	(1.2)	(0.6)	(1.2)	(1.9)
	24	37	9	7	13	10	13	12	8
pH[2]	7.3	7.3	7.3	7.3	7.3	7.2	7.3	6.8	7.3
	(0.1)	(0.3)	(0.1)	(0.1)	(0.1)	(0.3)	(0.1)	(0.2)	(0.5)
	24	39	8	8	13	11	14	12	8
Dissolved Oxygen in Mixed Liquor (mg/l)	6.0	6.6	5.6	5.2	4.2	4.8	8.4	6.4	6.0
	(1.9)	(0.4)	(0.3)	(0.5)	(0.4)	(2.3)	(0.4)	(0.7)	(0.1)
	13	16	8	6	11	9	10	12	2
Dissolved Oxygen in Dilution Water (mg/l)	1.6	2.2	2.2	1.5	3.1	3.3	3.9	1.4	3.1
	(0.9)	(1.8)	(1.8)	(0.5)	(1.4)	(2.7)	(4.0)	(0.5)	(2.4)
	12	16	8	6	10	9	9	12	2
Hydraulic Loading ($l/d \cdot m^2$)	71	70	71	70	70	72	71	73	68
	(3)	(9)	(3)	(2)	(5)	(2)	(4)	(2)	(18)
	16	16	9	7	11	10	7	11	8
HRT (min)	119	121	119	120	120	117	120	116	124
	(6)	(15)	(5)	(3)	(8)	(3)	(7)	(3)	(33)
	16	16	9	7	11	10	7	11	8

LEGEND: The first number in each set of three is the mean, the second is the standard deviation and the third is the number of data points.

1 - Data from Cruz (1977)

2 - pH is the arithmetic mean

be due to the stripping of carbon dioxide from the RBC mixed liquor at a rate faster than the heterotrophic bacteria can produce the carbon dioxide.

The mean mixed liquor dissolved oxygen concentration during the experiments was 6.0 mg/l or higher until the influent COD concentration was increased above 90 mg/l in phases III, IV, and V. As the influent COD concentration increased (see Table 5), the dissolved oxygen in the mixed liquor decreased, indicating that dissolved oxygen utilization increased. The decrease in the mixed liquor dissolved oxygen during phase V may have been larger if the dissolved oxygen level in the dilution water had not increased (due to a malfunctioning regulator on the nitrogen gas cylinder). The mean hydraulic loading on the RBC during the experiments varied between 68 $1/d \cdot m^2$ and 73 $1/d \cdot m^2$. The variation in the mean hydraulic loading between the experiments was insignificant and was approximately constant throughout the study. The hydraulic retention time, which is a function of hydraulic loading, was also constant for all of the experiments.

The disc surface area for biological growth was constant throughout each experiment. However, the wastewater on the submerged area of the discs (1.0 m^2) would sometimes flow

over the non-submerged portion of the disc (0.20 m^2), as the discs rotated, allowing a small amount of biomass to grow on this non-submerged portion. This therefore increased the surface area available for biological growth to as much as 1.2 m^2; however, the growth on this non-submerged portion was not consistently present. Hence, only the submerged surface area (1.0 m^2) was used in calculations of hydraulic and organic loading rates.

The data presented in Table 6 were from the steady state period during each experiment, with the exception of phase V. During phases VA and VB, the RBC was in a state of transition throughout both experiments. The selection of the appropriate periods for analysis are discussed in the following section.

Steady State Operation

The RBC unit was operated continuously from September 10, 1976 (Julian calendar day 253) when Cruz (1977) began his experiments to September 25, 1977 (Julian calendar day 281). The only significant changes in the operation of the RBC system during this interval were θ_{op}, influent COD concentration and influent nitrogen source. These three control parameters are presented in Table 7 along with the influent nitrogen concentrations, the duration of each experiment and the steady state time periods. The selection of steady state

Table 7

Control Parameters

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
Operating Time Period[2]	100-148	149-207	208-243	244-256	257-269	270-280	18-55	61-98	251-329
State	steady	steady	steady	steady	transit	transit	steady	steady	steady
Time Period	124-148	166-207	235-243	248-256	257-269	270-280	42-55	86-98	316-329
θ_{op} (days)	18	18	12	12	12	12	6	12	—[3]
Influent COD	62	84	90	136	172	190	57	63	60
COD	(5)	(17)	(6)	(5)	(20)	(25)	(6)	(3)	(4)
(mg/l)	9	15	9	7	11	10	13	7	14
Nitrogen Source	Ammonia	Glycine	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia
Influent	15.6	23.1	17.5	20.5	20.2	18.0	18.8	19.8	19.6
Nitrogen	(2.1)	(2.4)	(1.2)	(1.3)	(1.8)	(1.3)	(3.2)	(0.9)	(0.6)
(mg/l)	7	10	9	6	6	3	14	7	13

LEGEND: The first number in each set of three is the mean, the second is the standard deviation and the third is the number of data points.

1 - Data from Cruz (1977)

2 - Time is in Julian calendar days

3 - No discs were scraped during run 3

time periods for runs 1, 2 and 3 are presented by Cruz (1977). For clarification and continuity of operation, the order in which the various experiments were performed was run 3, run 1, run 2, and phase I through VB.

Phase I

The influent and effluent COD concentrations during phase I are shown in Figure 13. Phase I was started immediately after run 2; the only change was the increase in θ_{op} from 12 d to 18 d. The average influent COD concentration during the latter part of phase I was 62 mg/l; this concentration was assumed to be approximately the same during the first part of this phase. This assumption is based on the fact that the RBC system had been operating for more than 200 d prior to phase I with an approximate influent COD concentration of 60 mg/l and the preparation of the organic substrate had not been changed between run 2 and phase I. The effluent COD concentration generally responded to fluctuations in the influent COD concentration; otherwise, the effluent COD concentration was relatively constant.

The concentration of the influent and effluent nitrogen species are presented in Figure 14. On day 114, the nitrate-nitrogen concentration decreased to 1 mg/l and 5 d later returned to the concentration prior to the decrease. The mean

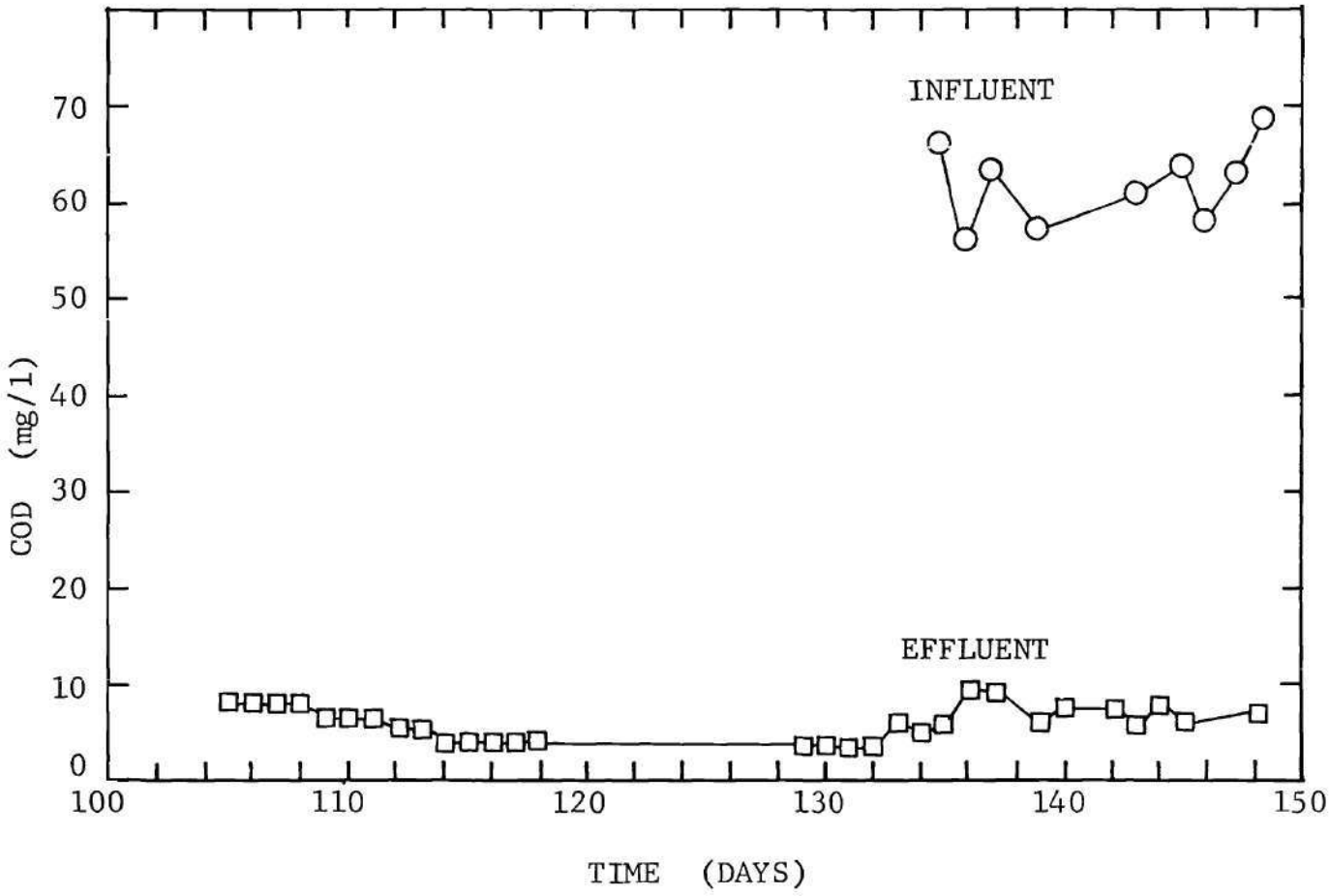

Figure 13. RBC Influent and Effluent COD Concentrations During Phase I

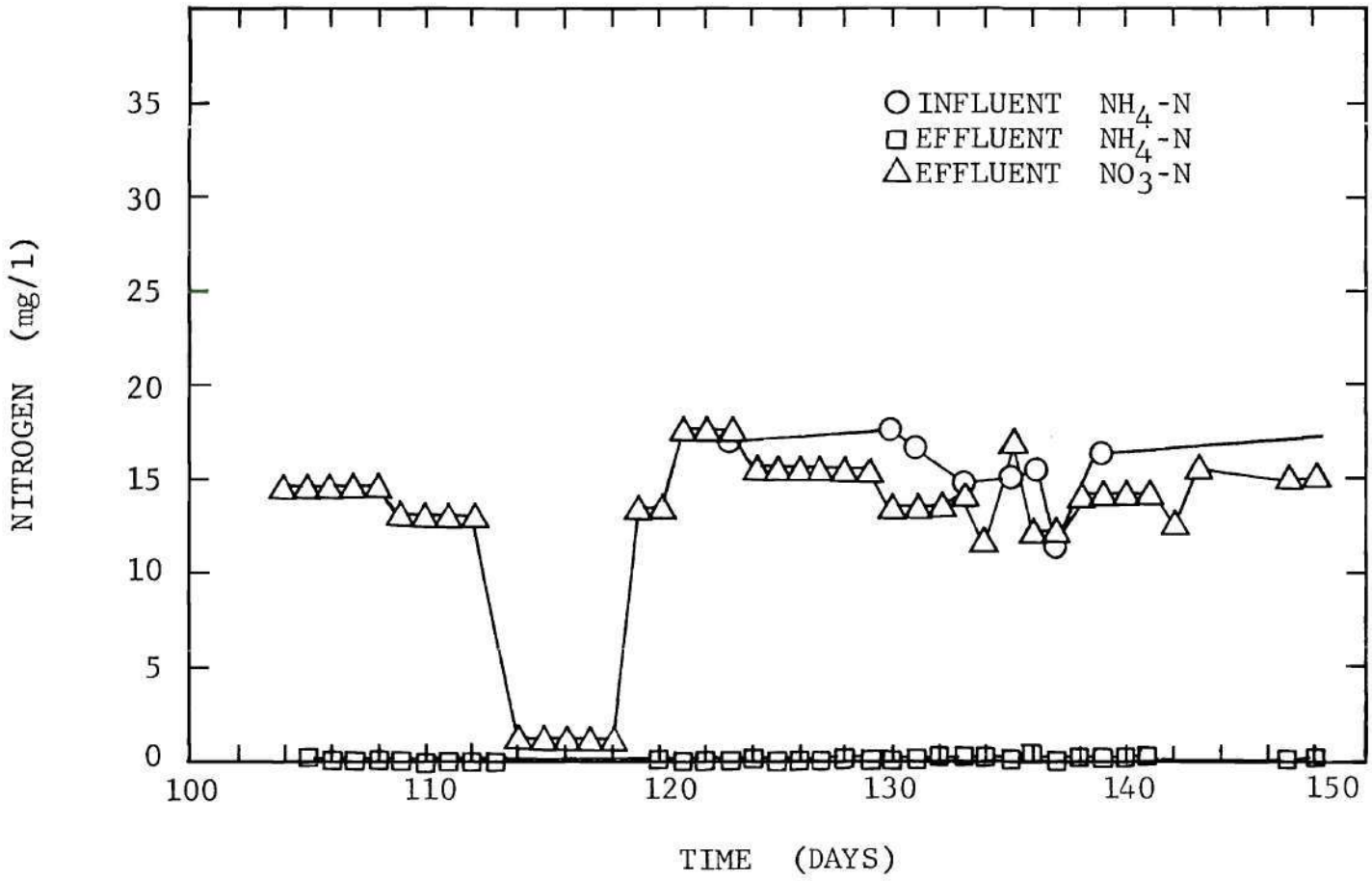

Figure 14. RBC Influent and Effluent Nitrogen Species Concentrations During Phase I

effluent nitrate-nitrogen concentration was 14.3 mg/1 during the remainder of phase I. This concentration did not vary significantly throughout the following phase, indicating that nitrification was at steady state during the last half of phase I. Also, to support this conclusion was the fact that the effluent ammonia-nitrogen concentration was very low throughout this phase. The reason for the decrease in the nitrate-nitrogen concentration cannot be determined; however, a possible reason may be that denitrification was occurring in the sample bottles during storage. The RBC system was considered to be at steady state from day 124 to day 148.

Phase II

For phase II, the nitrogen source was changed from ammonia to glycine. The influent COD concentration was increased a small amount due to the addition of glycine as a nitrogen source. The θ_{op} value was continued as 18 d. The influent and effluent COD concentrations during phase II are shown in Figure 15. The effluent COD concentration initially increased until the microorganisms could acclimate and metabolize the glycine. The influent COD concentration gradually increased during the phase; however, the effluent COD concentration remained relatively constant with only minor fluctuations.

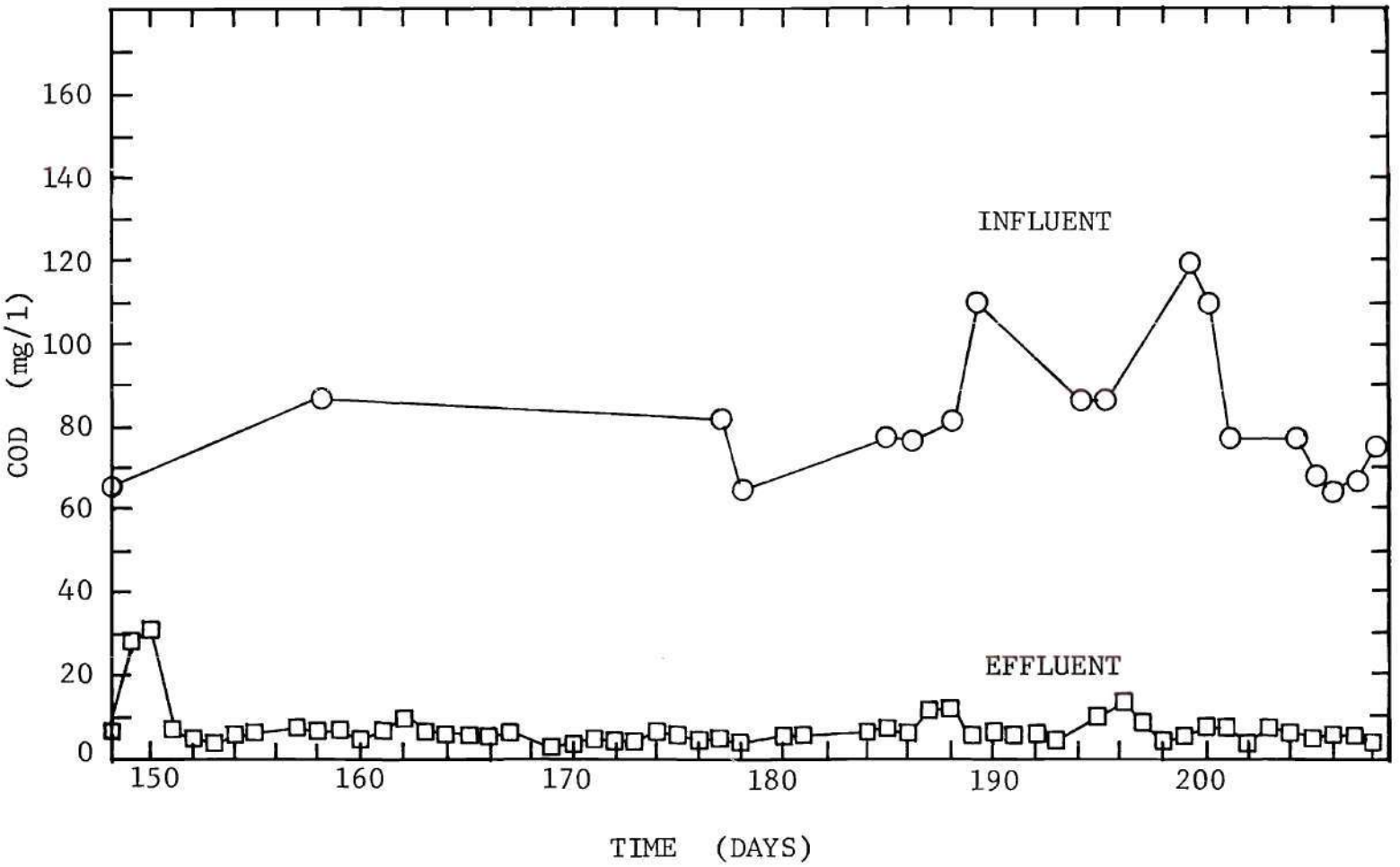

Figure 15. RBC Influent and Effluent COD Concentrations During Phase II

The influent glycine and effluent ammonia-nitrogen concentrations were also relatively constant with minor fluctuations as shown in Figure 16. The effluent nitrate-nitrogen concentration fluctuated; however, these fluctuations could not be correlated to any operating condition, e.g. scraping frequency. Since the effluent nitrate-nitrogen concentration did not drift significantly, the RBC system was considered to be at steady state between day 166 and day 207.

Phase III

During phase III, the sole nitrogen source was ammonia-nitrogen and the θ_{op} was decreased from 18 d to 12 d. The influent COD concentration was increased as shown in Figure 17. The effluent COD concentration increased slightly; afterwards, the influent and effluent COD concentrations were constant.

The influent and effluent nitrogen species are presented in Figure 18. The initial high influent ammonia-nitrogen concentration resulted from an error in preparing the inorganic substrate and was corrected the following day. The effluent ammonia-nitrogen and nitrate-nitrogen concentrations fluctuated during the major portion of phase III. However, toward the end of phase III, the influent and effluent ammonia-nitrogen and effluent nitrate-nitrogen concentrations were

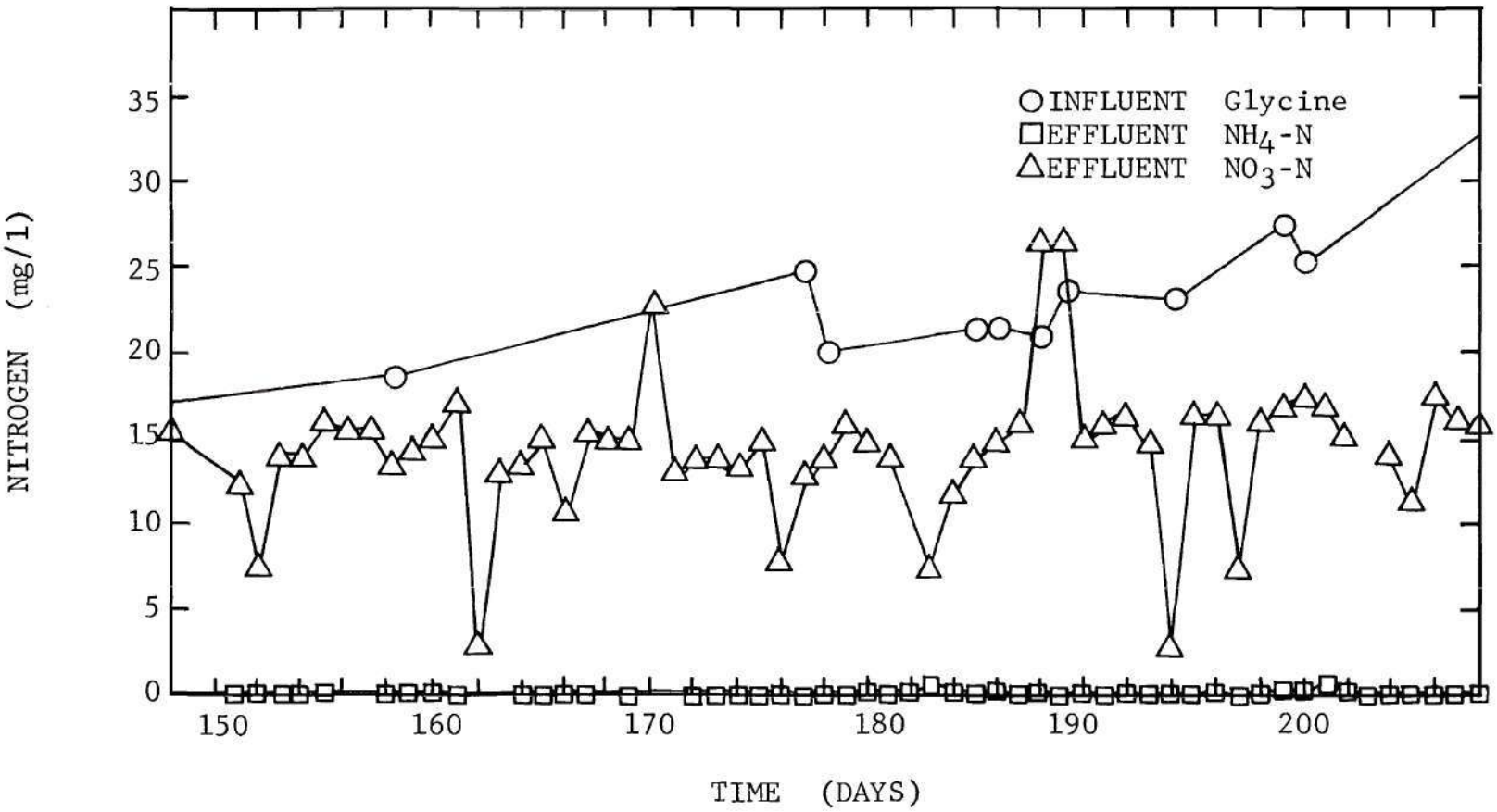

Figure 16. RBC Influent and Effluent Nitrogen Species Concentrations During Phase II

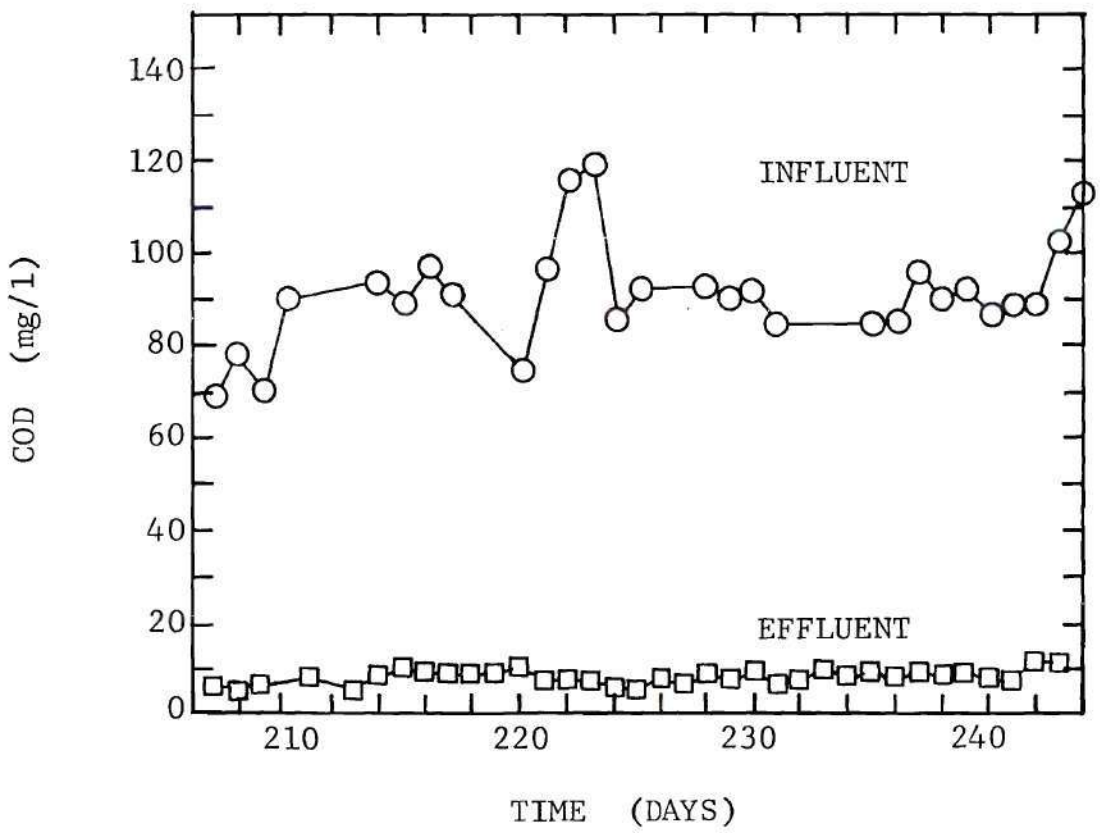

Figure 17. RBC Influent and Effluent COD Concentrations During Phase III

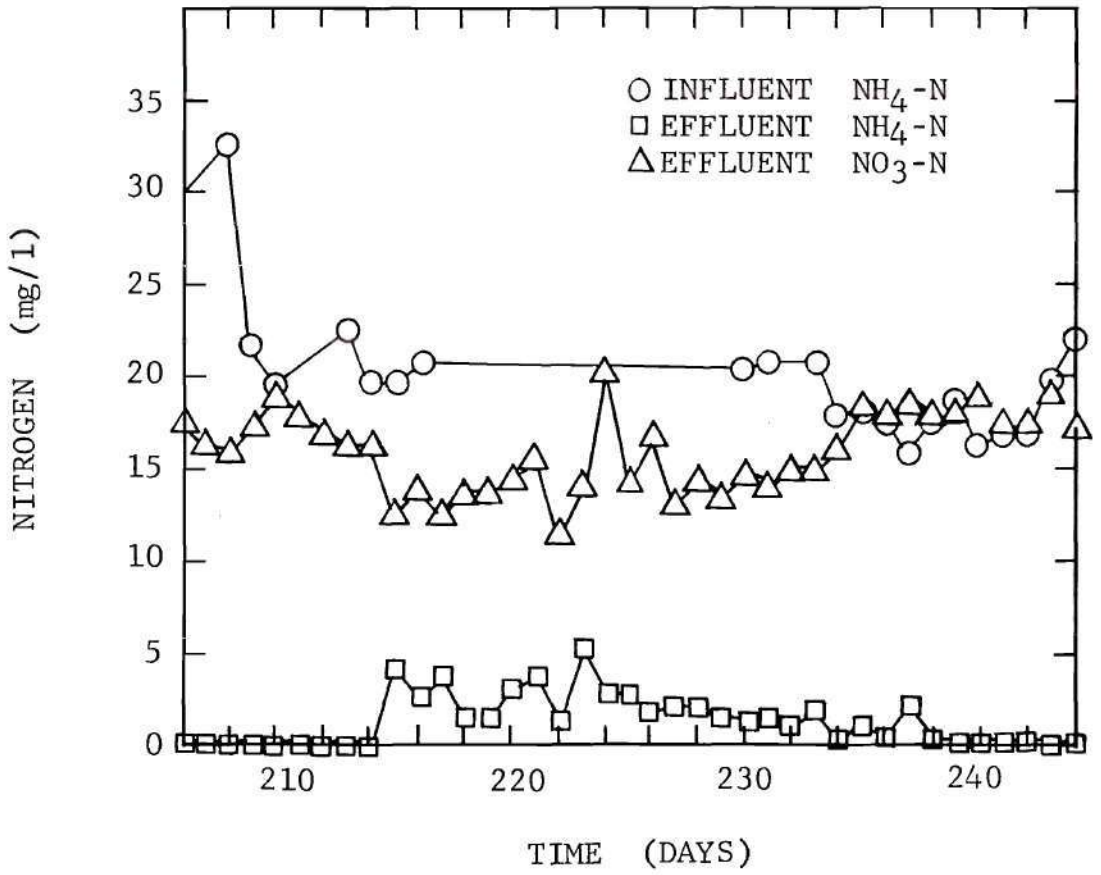

Figure 18. RBC Influent and Effluent Nitrogen Species During Phase III

approximately constant; during this time period, day 235 to day 243, the RBC system was considered to be at steady state.

Phase IV

During phase IV, the θ_{op} remained at 12 d and the influent ammonia-nitrogen concentration remained approximately the same as in phase III. The influent COD concentration was increased by approximately 40 mg/l over a 3 d period to approximately 140 mg/l as shown in Figure 19. The effluent COD concentration remained constant and did not reflect the increase in the influent COD. The effluent ammonia-nitrogen concentration was constant throughout the phase, as shown in Figure 20. Fluctuations did occur in the influent ammonia-nitrogen and effluent nitrate-nitrogen concentrations; however, they were relatively constant during the latter part of the phase. The RBC reactor was considered to be at steady-state from day 248 to day 256.

Phase V

The θ_{op} during phase V was 12 d and the influent ammonia-nitrogen concentration was not changed significantly from phase IV. Phase V was subdivided into two parts, A and B. The influent COD concentration during part A was increased by approximately 35 mg/l and held at approximately 170 mg/l as shown in Figure 19. The effluent COD and ammonia-nitrogen

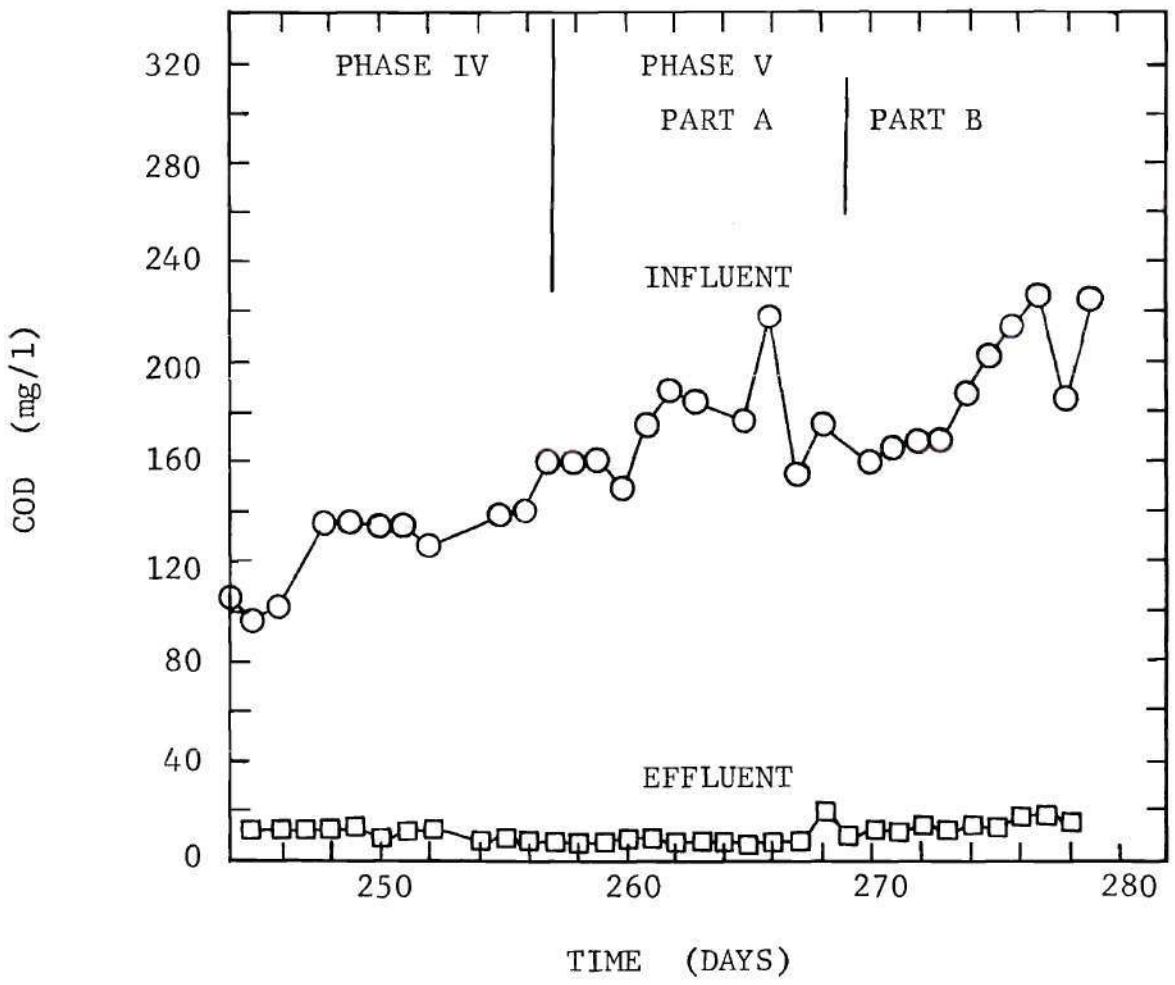

Figure 19. RBC Influent and Effluent COD Concentrations During Phases IV and V

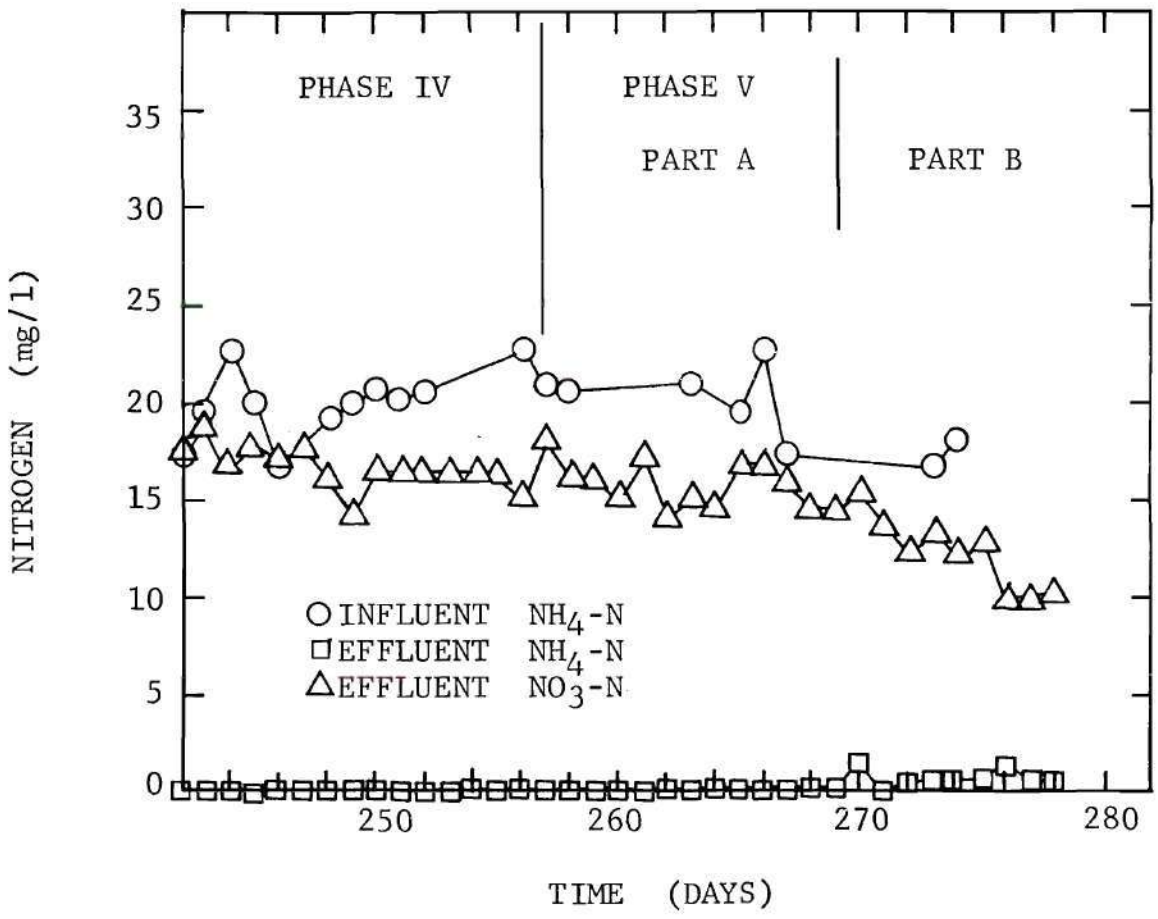

Figure 20. RBC Influent and Effluent Nitrogen Species During Phases IV and V

concentrations were relatively constant at 9 mg/l and 0.21 mg/l, respectively. The influent ammonia-nitrogen and effluent nitrate-nitrogen fluctuated throughout part A. The RBC was not considered to be at steady state; however, the data collected during part A, day 257 to day 269, was averaged to provide some general information on the performance during this time.

The influent COD concentration was increased by approximately 8 mg/l per day during the first 8 d of phase VB and held at approximately 200 mg/l for the last three days, as shown in Figure 19. The effluent COD concentration gradually increased during this period from 11 mg/l to 17 mg/l. The influent and effluent ammmonia-nitrogen concentrations increased during part B and the effluent nitrate-nitrogen concentration decreased. Steady state was not achieved during this part of phase V. However, as in part A, the data collected from day 270 to day 280 were averaged to provide some general information on performance. The RBC system during phase V was considered to be in a transitory state.

Cruz (1977) performed three experiments (i.e. runs 1, 2 and 3) using the same RBC system. The influent COD and ammonia-nitrogen concentrations were approximately constant during all three runs. The θ_{op} values were 6 d and 12 d for

runs 1 and 2. The discs were not scraped during run 3. The RBC system was assumed to be at steady state when the effluent COD, ammonia-nitrogen and nitrate-nitrogen concentrations were relatively constant. The data collected during the steady state periods were averaged to determine the performance of the RBC system.

Solids data during phase I through V were collected as well as COD, ammonia-nitrogen and nitrate-nitrogen data. The biomass, measured as TSS, scraped from the discs every $\theta_{op}/6$ d was plotted and is shown in Figure 21. The biomass on these discs fluctuated considerably during the 5 phases. Similar results were also observed by Cruz (1977). The general changes in biomass will be discussed later in more detail. These fluctuations were due to random sloughing of the biofilm from the discs. Hence, the biomass in the RBC system was always fluctuating; therefore, the system was never at a true steady state. However, since the oxidation of the COD and ammonia did reach steady conditions, the RBC was assumed to be at a pseudo steady state.

Cell Residence Time

To determine cell residence time (θ_c), a portion of biofilm was scraped from the discs at the end of phases I, II and III to obtain a biomass distribution. The biomass per

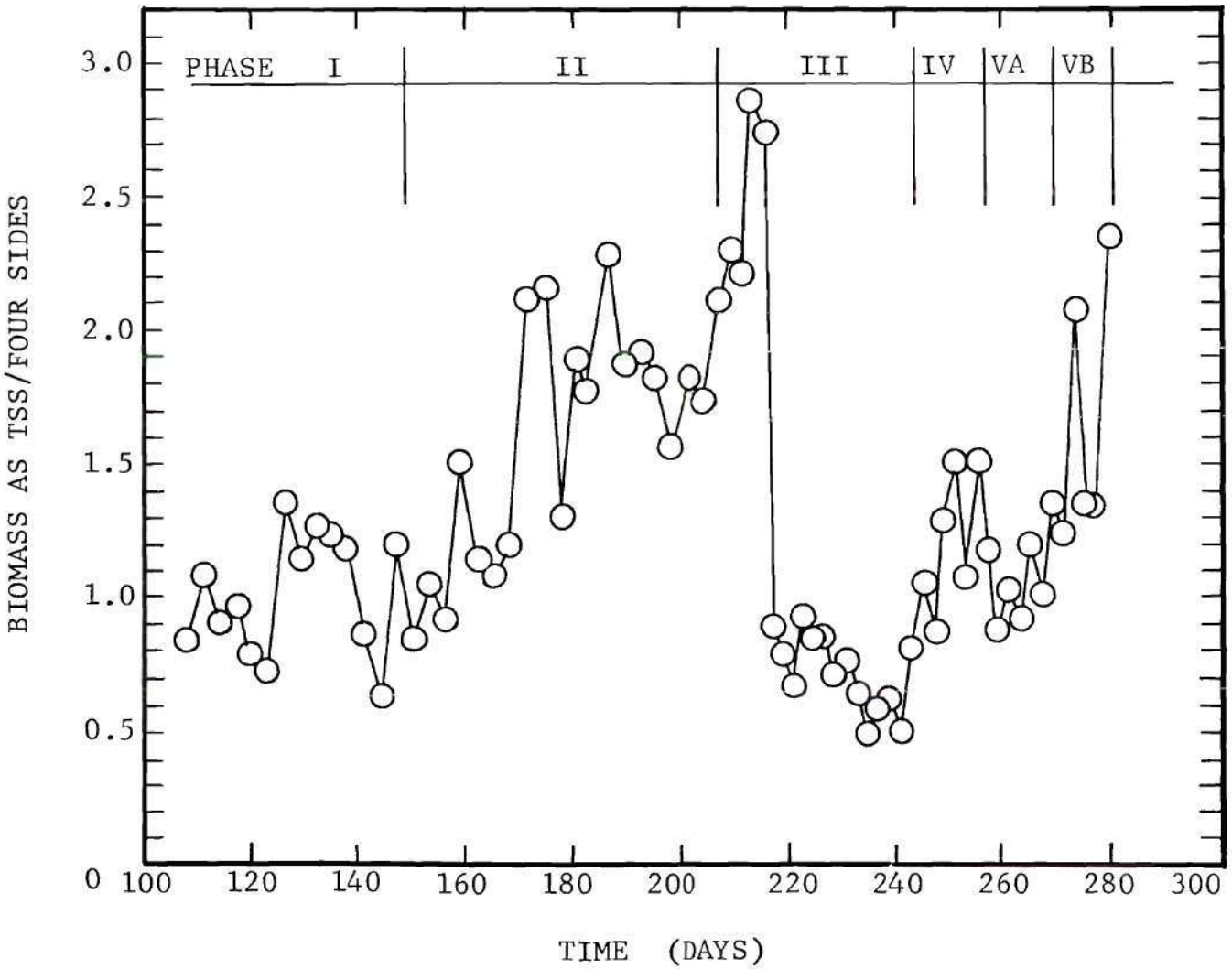

Figure 21. Biomass (as TSS) Scraped From Discs

disc side was plotted against age (the time elapsed since the side was last scraped) and are shown in Figures 22, 23 and 24 for phases I, II and III, respectively. Similar figures were presented by Cruz (1977). The biomass distribution was not determined during phases IV and V. The biomass per disc side generally increased with age; however, no particular function fit all the distribution curves. The biomass on each of the disc sides was assumed to have fluctuated as was observed in Figure 21. Since the biomass distribution curves would be continually changing, a linear distribution of the biomass with age was assumed to determine an approximate average biomass on the discs. As shown in Figures 22 and 23, linear relationships sould be used to fit the data; however, a linear relationship could not effectively be used to fit the data for phase III as shown in Figure 24.

The solid data points in Figures 22, 23 and 24 represent the average biomass per disc side which was scraped from the discs every $\theta_{op}/6$ d during the respective steady state periods. The solid data points which represented the average value for many data points generally did not correspond to the biomass distribution curves which were instantaneous biomass measurements. Therefore, in calculating the biomass on the discs, it was assumed that the biomass was linearly

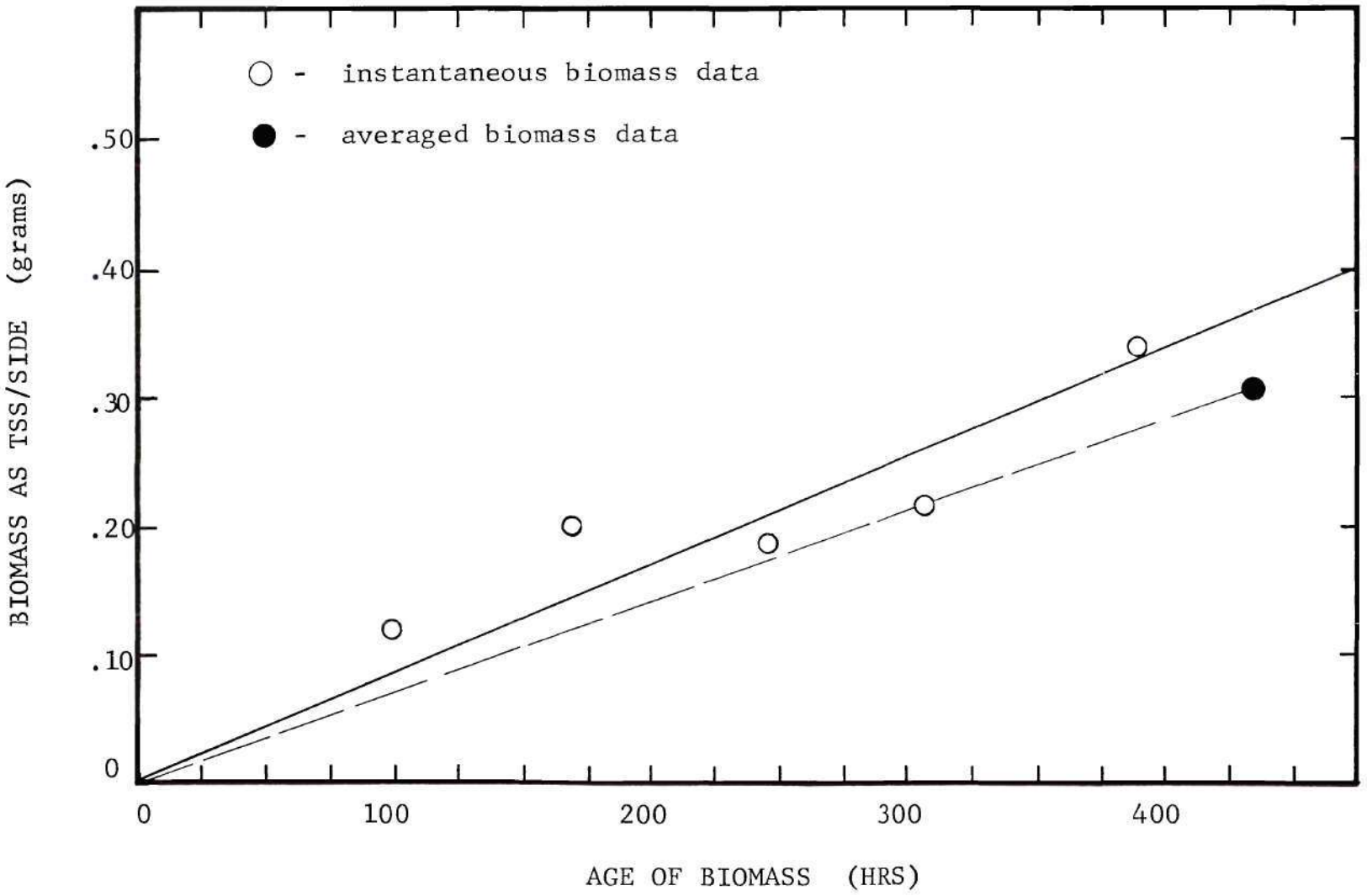

Figure 22. Biomass Distribution on the Disc Sides During Phase I

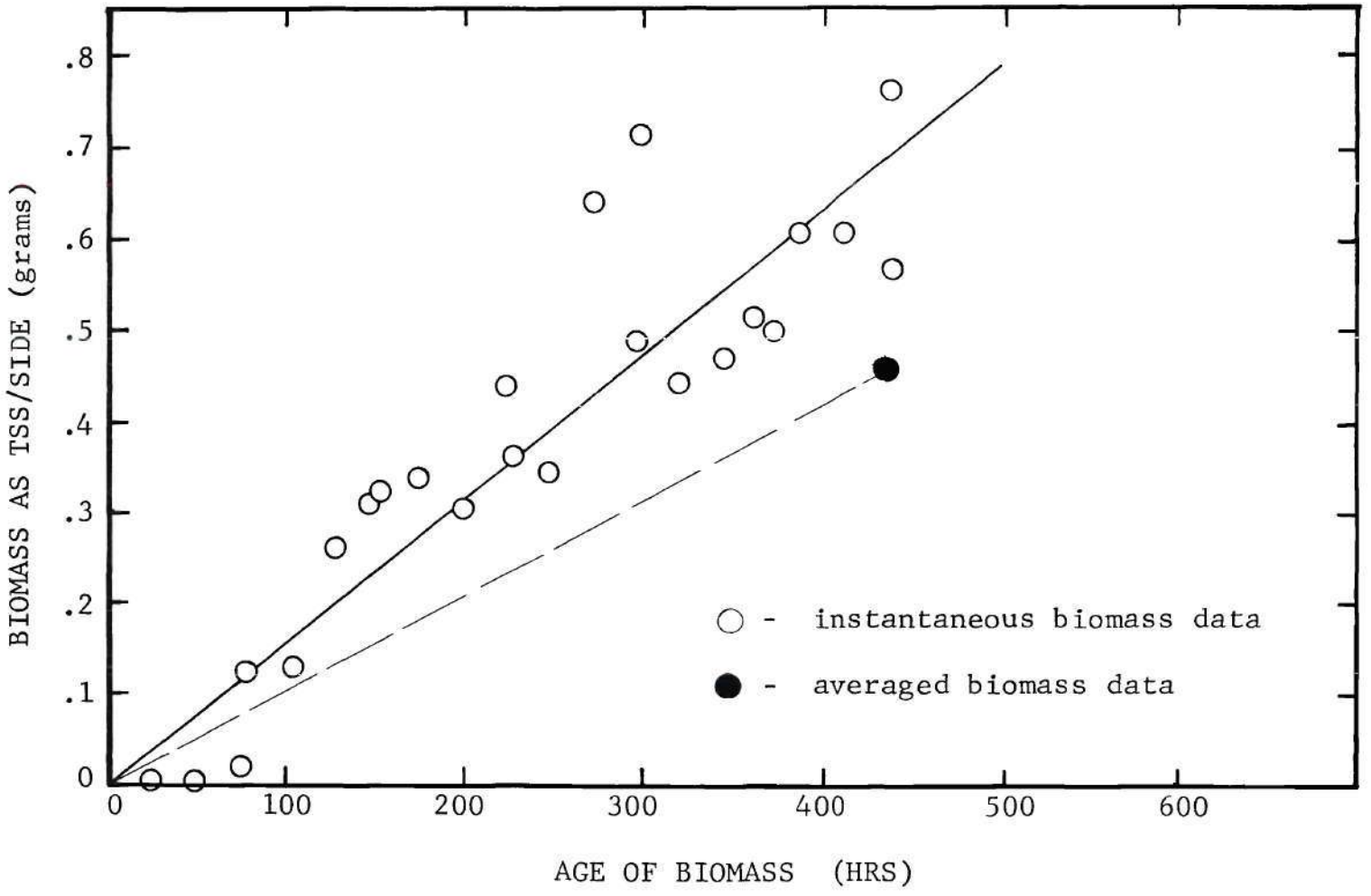

Figure 23. Biomass Distribution on the Disc Sides During Phase II

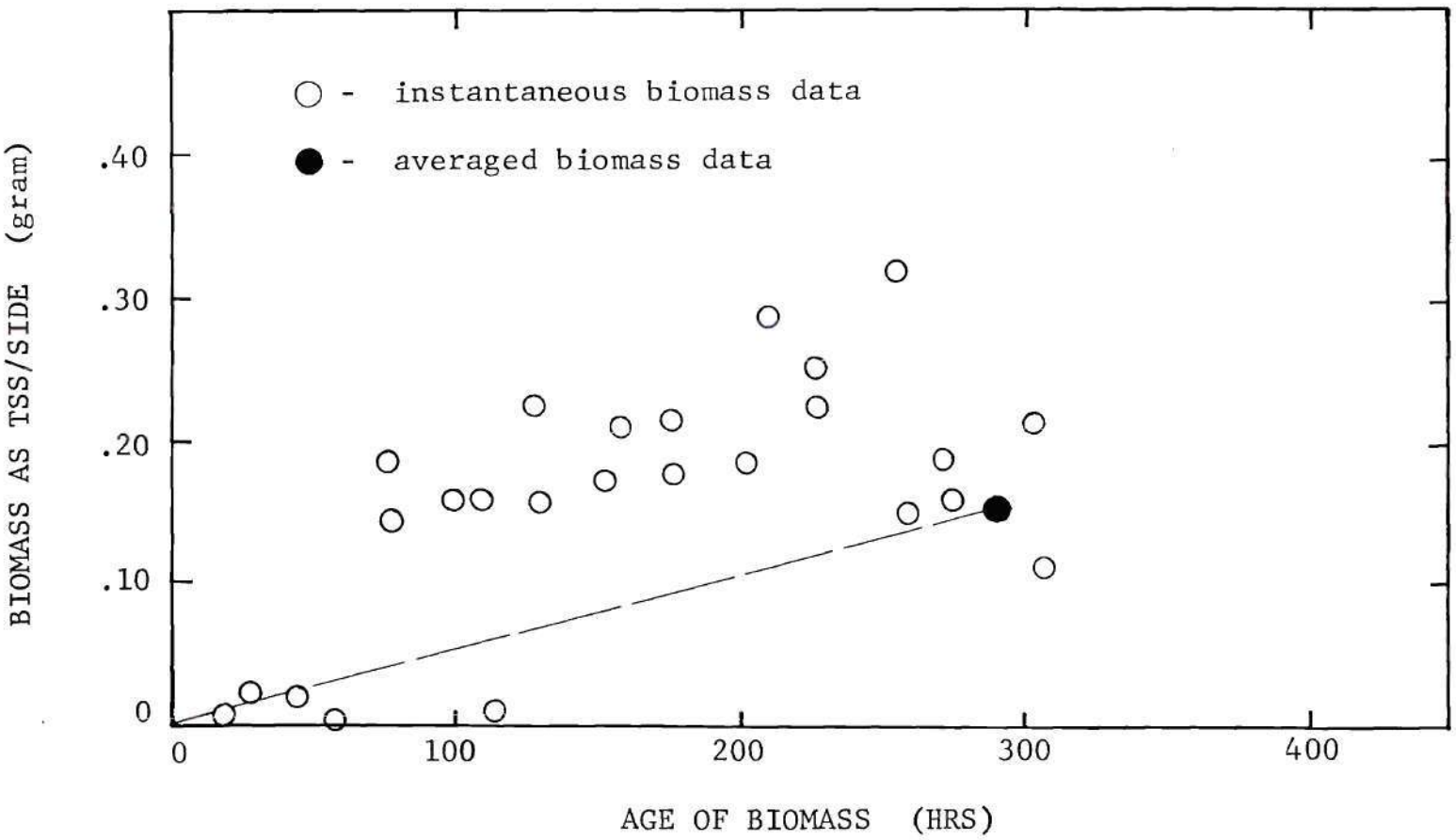

Figure 24. Biomass Distribution on the Disc Sides During Phase III

distributed between zero and the average biomass scraped from the discs every $\theta_{op}/6$ days, as shown by the dotted lines in Figures 22, 23 and 24.

The biomass on the discs was calculated using the following method. The mass collected (measured as TSS) from the 4 sides scraped every $\theta_{op}/6$ d (SS_D) was averaged and expressed as $\overline{K}$. The total mass on the discs just prior to scraping (M_{D1}) was then:

$$M_{D1} == \overline{K} \left(\frac{\sum_{i=1}^{6} i}{6} \right) \qquad (10)$$

The total mass on the discs just after scraping (M_{D2}), assuming the scraped discs had zero mass, would be:

$$M_{D2} = \overline{K} \left(\frac{\sum_{i=1}^{6} i}{6} \right) \qquad (11)$$

The average mass on the discs (M_D) would then be the average of equations (10) and (11) and would be:

$$M_D = 3\,\overline{K} \qquad (12)$$

The majority of the suspended solids in the mixed liquor appeared to have sloughed from the discs. These suspended solids were in large clumps and had the same visual appearance as the attached biomass. As a result of the rotation of the discs, shearing forces were exerted on the biomass as it passed through the wastewater and some biomass was stripped from the disc surface into the mixed liquor. These sloughed solids were either continuously removed with the sieves in the circulation system (SS_c) or sieved from the mixed liquor every $\theta_{op}/6$ d (SS_M). The biomass (measured as TSS) in the mixed liquor (M_{ML}) was calculated by averaging the sum of the solids fractions SS_M and SS_C. This approach is based on the assumption that the MLSS were equal to zero initially after the SS_M and SS_c were collected and that the MLSS increased linearly until collected again. The total mass in the system (M) was then the sum of M_D and M_{ML}:

$$M = 3\overline{K} + M_{ML} \qquad (13)$$

The total biomass wasted (R_w), measured as TSS, was the average of the sum of SS_D, SS_E, SS_M and SS_c (see the Appendix for the actual values for these solids fractions). The cell residence time (θ_c) was calculated by dividing the total

biomass (M) by the biomass wasted (R_w):

$$\theta_c = \frac{M}{R_w} \tag{14}$$

The values for R_w, M_{ML}, M_o and θ_c for each experiment are presented in Table 8. The solids data from run 3 (Cruz, 1977) were insufficient for use in calculating θ_c. Phases III, IV, VA and VB and run 2 had θ_{op} values of 12 d and have θ_c values that are similar even though the M and R_w values were varied. The average of these 5 θ_c values (i.e. 2 d) will be used as the designated θ_c value for these 5 experiments. For phases I and II, the average θ_c (i.e. 3.5 d) will also be used. For run 1, the calculated θ_c value of 1.3 d will be used. The present by Cruz (1977) was used in the calculation of θ_c values for runs 1, 2 and 3. However, his calculation procedure was not used; therefore, the θ_c values do not agree. Cruz (1977) in his θ_c calculations did not take into account the biomass suspended in the mixed liquor as part of the total mass in the system.

A factor that may have influenced θ_c was the active biofilm depth. The biofilms on the RBC discs were approximately 1 mm in thickness, i.e. 1000 μm. From the data

Table 8

Cell Residence Time Data

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3[2]
θ_{op} (days)	18	18	12	12	12	12	6	12	—
M_{ML} (gm)	1.45	1.74	1.19	2.18	1.98	2.92	0.20	1.27	—
M_D (gm)	3.32	5.49	1.85	4.07	3.29	5.04	1.43	4.37	—
M (gm)	4.77	7.23	3.04	6.25	5.27	7.96	1.63	5.64	—
R_w (gm/d)	1.43	2.03	1.66	3.18	2.66	4.04	1.26	2.4	—
θ_c (days)	3.3	3.6	1.8	2.0	2.0	2.0	1.3	2.4	—
θ_c[3] (days)	3.5	3.5	2.0	2.0	2.0	2.0	1.3	2.0	—

LEGEND:

1 - Data from Cruz (1977)

2 - Solids data was insufficient to calculate θ_c

3 - The average θ_c, with respect to the θ_{op}. These θ_c values will be used for the experiments.

collected by others and presented earlier in Table 1, active biofilm depths ranged from 37 to 200 μm. Based on this information, the biofilm depth measured on the RBC discs had a considerable thickness of non-active biofilm. However, if active biofilm depths had been determined for this study and were used, the biomass (M) would have been decreased; however, the wastage rate would have been decreased proportionately and the θ_c probably would not have been changed significantly.

Performance

Organic Removal

The RBC system removed soluble organic matter, measured as COD, and oxidized ammonia to nitrate; each of these capabilities will be discussed individually. Some parameters used to evaluate the performance of the RBC in removing organic matter are presented in Table 9. These values are averages taken during the steady state or evaluation periods indicated earlier. For influent and effluent COD concentrations, the standard deviation and number of data points used are also listed. The effluent COD concentration was practically the same for all the experiments, with the exception of phase VB, which only had a relatively small increase in effluent COD concentration. The RBC system was capable of receiving an influent COD concentration up to 190 mg/l without any major

Table 9

Influent and Effluent COD Concentrations

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
θ_c (days)	3.5	3.5	2.0	2.0	2.0	2.0	1.3	2.0	—
Influent COD (mg/1)	62	84	90	136	172	190	57	63	60
	(5)	(17)	(6)	(5)	(20)	(25)	(6)	(3)	(4)
	9	15	9	7	11	10	13	7	14
Effluent COD (mg/1)	6	7	10	11	9	16	10	7	7
	(2)	(3)	(4)	(2)	(4)	(2)	(2)	(2)	(3)
	16	37	9	8	13	9	13	7	14
COD Removal (%)	90	92	89	92	95	92	82	89	88

LEGEND: The first number in each set of three is the mean, the second is the standard deviation and the third is the number of data points.

1 - Data from Cruz (1977)

effects in effluent COD concentration.

No other research using a RBC system to treat secondary effluents has been found by the author; however, several cases are available for comparison with RBC systems treating raw wastewater or primary effluent. These cases have higher influent organic concentrations with respect to this study; therefore, phase VB will be used for comparison purposes even though the phase was not at steady state. In Table 10, the results reported by Hao and Hendricks (1975A), Torpey et al. (1971), Antonie (1976) and Smith (1976) are compared to phase VB. The BOD_5 concentrations used by these researchers were divided by 0.6 to obtain BOD_u values. The substrate used in this present study was completely biodegradable; therefore, the COD values are comparable to BOD_u values. The first three studies listed in Table 10 and Phase VB had influent BOD_u concentrations of approximately 200 mg/1, effluent BOD_u concentrations ranging from 11 mg/1 to 16 mg/1 and removals between 92% and 94%. The performance of all four of these RBC systems were relatively equal. However, comparing the results reported by Smith (1976) to phase VB at approximately the same treatment conditions, i.e. the hydraulic loadings and influent BOD_u concentrations were similar, it was observed that the secondary effluent could probably be treated

Table 10

Comparison of Influent and Effluent Organic Concentrations

With Other RBC Systems

	Hao and Hendricks (1975A)	Torpey, et al. (1971)	Antonie (1976)	Smith (1976)	PHASE V B[1]
Scale	Pilot	Pilot	Pilot	Bench	Bench
Wastewater	Settled Domestic and Industrial	Raw Domestic Wastewater	Settled Domestic	Shellfish Processing Wastewater	Secondary Effluent
Hydraulic Loading ($l/d \cdot m^2$)	61	71	71	81	72
BOD_5 (mg/l)					
Influent	112	124	125	151	—
Effluent	6.4	9	9	24	—
BOD_u (mg/l)[2]					
Influent	187	207	208	250	190
Effluent	11	15	14	40	16
% Removal	94	93	93	84	92

LEGEND: 1 - COD assumed to equal BOD_u

2 - $BOD_u = \frac{BOD_5}{0.6}$

— Data not collected

to lower BOD_u concentration than shellfish processing wastewater.

Additional performance parameters were influent and effluent mass flows and mass loading rates; these are listed in Table 11. The loading rate was calculated by dividing the influent COD mass flow in $g/d \cdot m^2$ by the total biomass per unit surface area in g/m^2. The total biomass per unit surface area was calculated by dividing the total biomass in the system, M (calculated in the Cell Residence Time section) by 1.0 m^2, the total surface area. The removal rate was the difference between the influent and effluent COD mass flows divided by the total biomass per unit surface area.

At approximately the same influent COD mass flow, the loading rate and effluent COD mass flow increased as θ_c decreased, as shown in Figure 25 and 26, respectively. For the same θ_c values, the loading rate and effluent COD mass flow generally increased as the influent COD mass flow increased. However, removal rate increased as loading rate increased as observed in Figure 27.

These same performance parameters from several other RBC systems are listed in Table 12. These are the same systems discussed with reference to Table 10. The influent and effluent COD mass flows, in the RBC system used by Hao

Table 11

Additional Performance Parameters of Organic Removal

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
θ_c (days)	3.5	3.5	2.0	2.0	2.0	2.0	1.3	2.0	—
Influent COD Mass Flow ($g/d \cdot m^2$)	4.4	5.9	6.4	9.5	12.0	13.7	3.9	4.9	4.5
Effluent COD Mass Flow ($g/d \cdot m^2$)	0.4	0.5	0.7	0.8	0.6	1.1	0.8	0.5	0.4
Loading Rate (day^{-1})	0.92	0.82	2.11	1.57	2.35	1.72	2.46	0.89	0.18
Removal Rate (day^{-1})	0.84	0.75	1.88	1.43	2.23	1.58	1.96	0.80	0.16

LEGEND: 1 - Data from Cruz (1977)

— Data not available

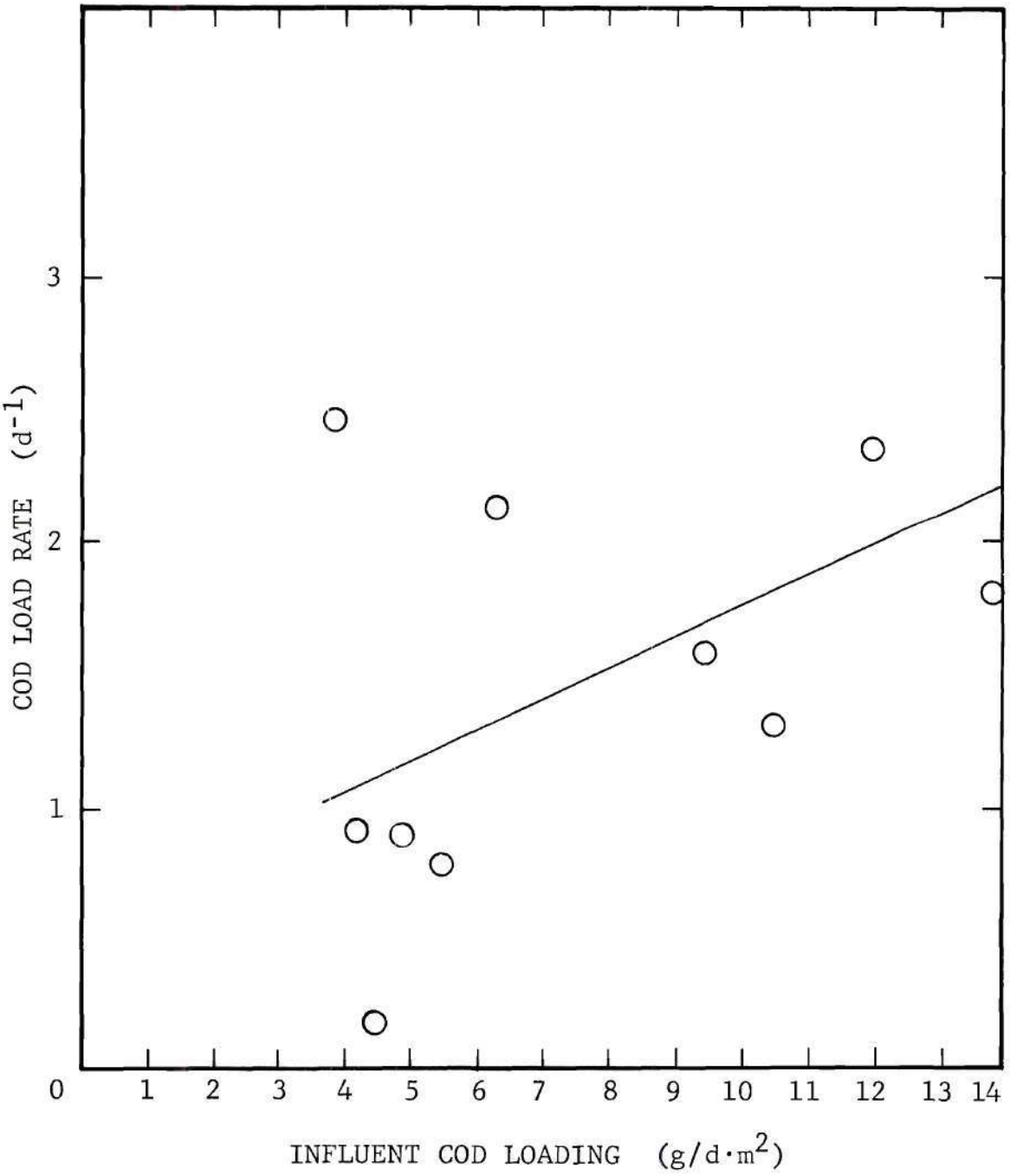

Figure 25. Influent COD Loading Vs. Loading Rate

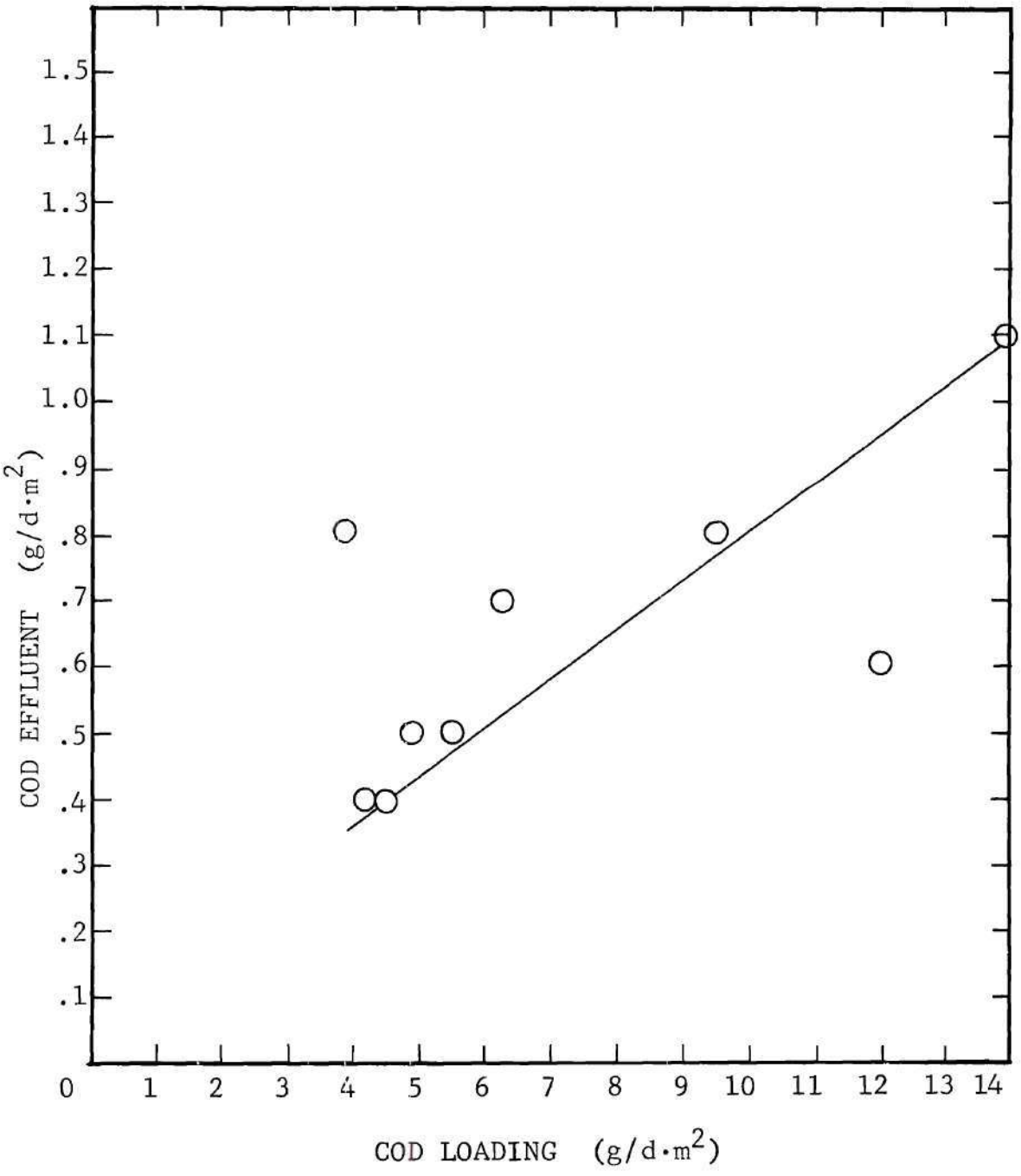

Figure 26. COD Loading Vs. COD Effluent

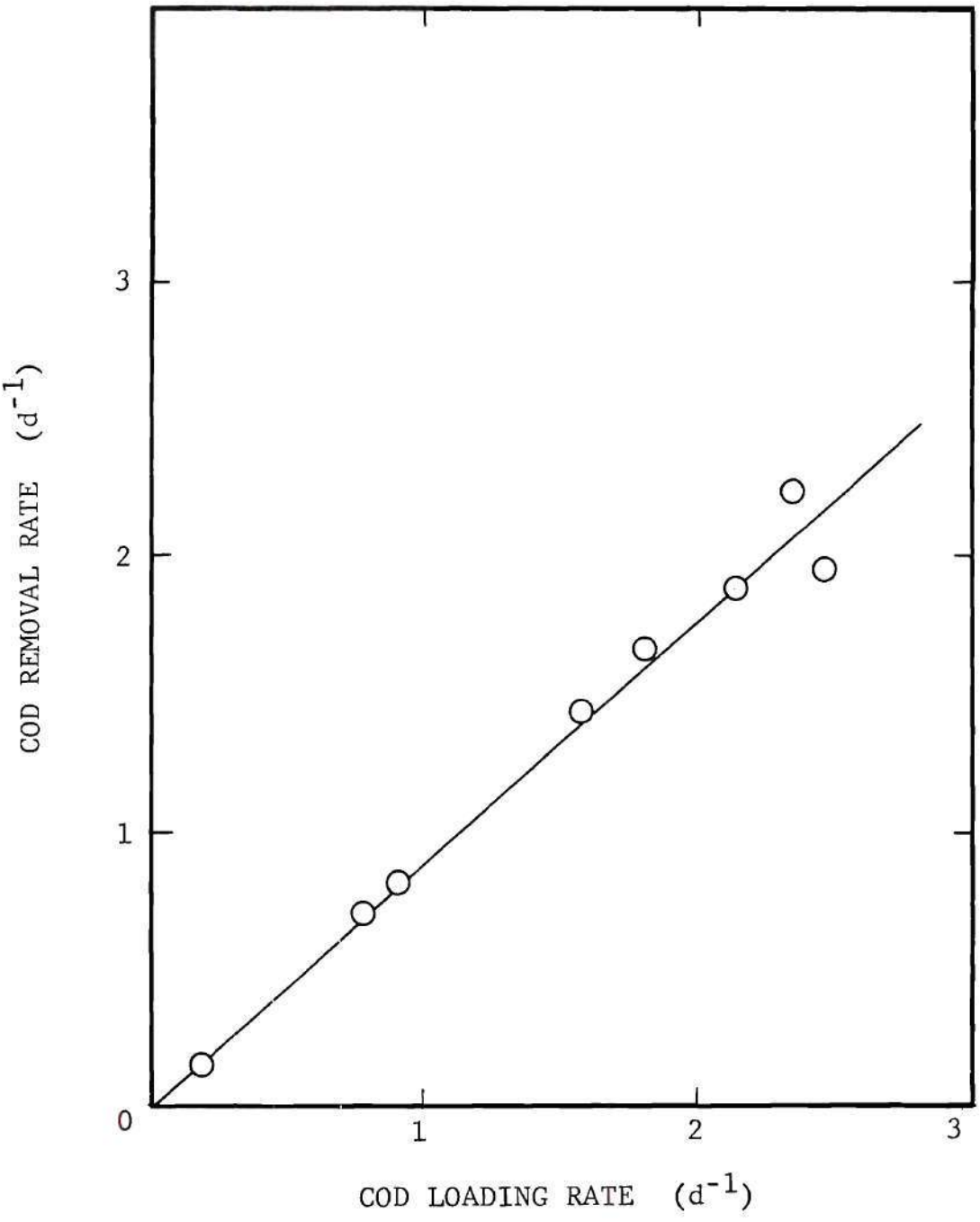

Figure 27. COD Loading Rate Vs. COD Removal Rate

Table 12

Comparison of Loading and Removal Rates
With Other RBC Systems

	Hao and Hendricks (1975A)	PHASE[1] V A	Torpey, et al. (1971)	Antonie (1976)	PHASE[1] V B	Smith (1976)	RUN[1] 3
BOD_u Mass Flow ($g/d \cdot m^2$)							
Influent	11.4	12.0	14.7	14.8	13.7	20.3	4.5
Effluent	0.7	0.6	1.1	1.0	1.1	3.2	0.4
Loading Rate (day^{-1})	—	2.35	—	—	1.72	0.31	0.18
Removal Rate (day^{-1})	—	2.23	—	—	1.58	0.26	0.16

LEGEND 1 - COD assumed to equal BOD_u

2 - $BOD_u = \frac{BOD_5}{0.6}$

— Data not available

and Hendricks (1975A), were approximately the same as in phase VA. The mass flows observed by Torpey et al. (1971) and Antonie (1976) are similar to those of phase VB. The high effluent BOD_u concentration reported by Smith (1976) can now be explained by the high influent BOD_u mass flow. The loading and removal rates used by Smith (1976) were similar to those in run 1; also, neither of these systems were scraped.

The RBC system used in this present study had comparable results in organic matter removal to other studies. The effluent COD concentration and COD percent removal generally increased with influent COD concentration with ranges from 6 to 16 mg/l and 82% to 95%, respectively. Performance was observed to improve with increasing θ_c values.

Nitrification

Some performance parameters that were used to evaluate the RBC system with respect to nitrification are listed in Table 13. The RBC system was able to remove glycine as well as ammonia, as demonstrated in phases II and I, respectively. The majority of the nitrogen from the glycine was completely oxidized to nitrate. No limitation of nitrification by high influent COD concentrations was observed. However, θ_c had a significant effect on nitrification as shown in Figure 28. These data also correspond to the results reported by many

Table 13

Influent and Effluent Nitrogen Species

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
θ_c (days)	3.5	3.5	2.0	2.0	2.0	2.0	1.3	2.0	—
Nitrogen Source	Ammonia	Glycine	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia	Ammonia
Influent Nitrogen (mg/l)	15.6 (2.1) 7	23.1 (2.4) 10	17.5 (1.2) 9	20.5 (1.3) 6	20.2 (1.8) 6	18.0 (1.3) 3	18.8 (3.2) 14	19.8 (0.9) 7	19.6 (0.6) 13
Effluent Nitrogen (mg/l)									
NH_4-N	0.17 (0.13) 20	0.28 (0.39) 39	0.48 (0.63) 9	0.12 (0.03) 6	0.21 (0.12) 13	0.76 (0.54) 9	14.9 (3.3) 14	1.03 (0.6) 7	0.10 (0.04) 13
NO_2-N	—	0.002 (0.001) 6	—	—	—	—	0.01 (0.004) 7	0.83 (0.21) 7	—
NO_3-N	14.3 (1.4) 22	14.8 (4.3) 40	16.7 (3.3) 9	16.0 (0.8) 9	15.9 (1.3) 13	12.3 (1.9) 9	0.17 (0.15) 14	18.7 (0.9) 7	16.1 (2.7) 6
TKN[2]	—	0.82 (0.66) 26	—	—	—	—	—	—	—
Ammonia-Nitrogen Removal (%)	99	99	97	99	99	96	21	95	99
Percent Nitrification (%)	99	93	97	99	99	94	0	91	99

LEGEND: 1 - Data from Cruz (1977)

2 - Soluble effluent TKN

— Data not collected

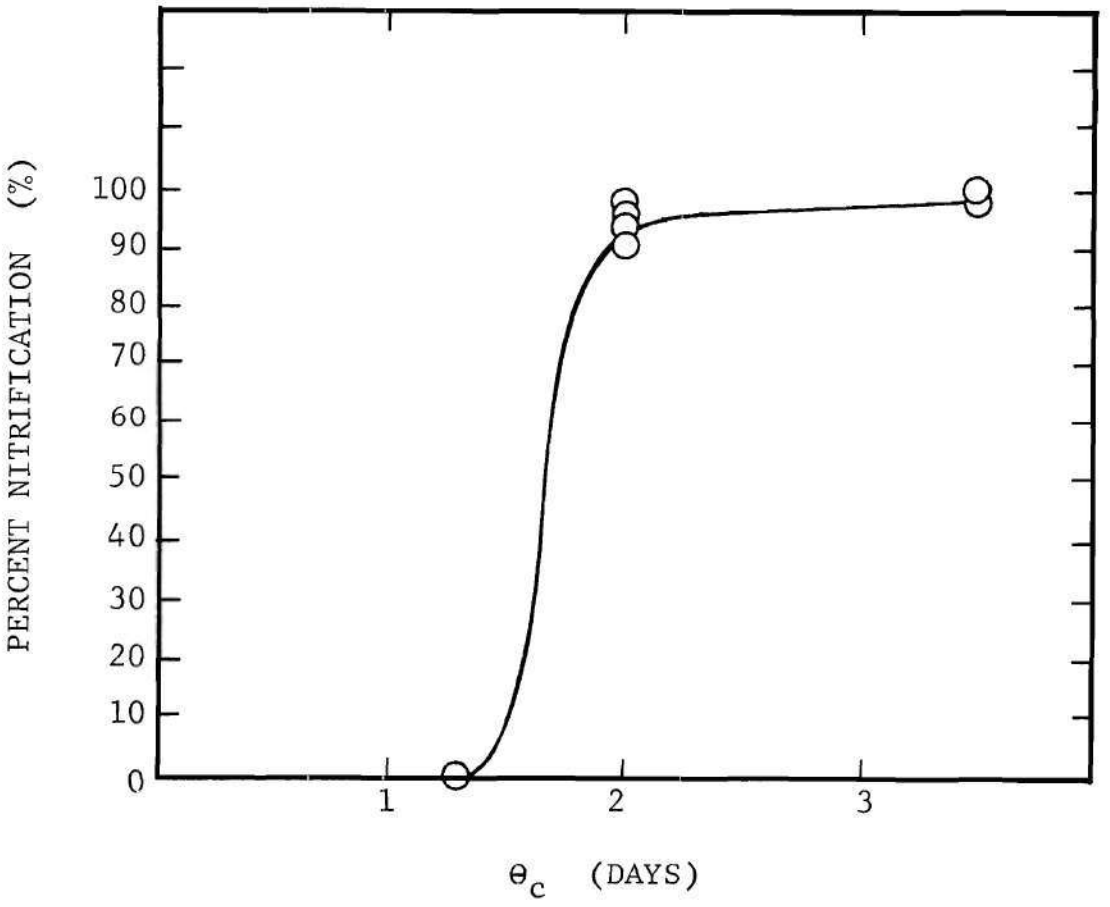

Figure 28. θ_c Vs. Percent Nitrification

others using different biological systems as shown in Figure 29. Based on the data compiled in Figure 29, the minimum θ_c value at which nitrification can occur was approximately 1.5 d. A θ_c value shorter than 1.5 d would not allow sufficient time for the retention of nitrifying microorganisms.

Several researchers have operated RBC systems under similar conditions to the system used in this study. Hao and Hendricks (1975A) used an RBC system with an influent COD and ammonia-nitrogen concentrations of 145 mg/1 and 11 mg/1, respectively, and a hydraulic loading of 61 $1/d \cdot m^2$. This RBC system achieved 92% ammonia removal. The RBC system in this study during phase III achieved a higher ammonia removal of 97% with influent COD and ammonia-nitrogen concentrations of 90 mg/1 and 17.5 mg/1, respectively and a hydraulic loading was 71 $1/d \cdot m^2$.

Antonie (1976) reported 97% ammonia-nitrogen removal with several RBC systems at influent COD, ammonia-nitrogen and TKN concentrations of 350 mg/1, 18 mg/1 and 29 mg/1, respectively. The hydraulic retention time was 120 min. Phase VB had similar conditions with a hydraulic retention time of 117 min and influent COD and ammonia-nitrogen concentrations of 190 mg/1 and 18 mg/1, respectively. During this phase, 96% ammonia-nitrogen removal was achieved. The RBC system in

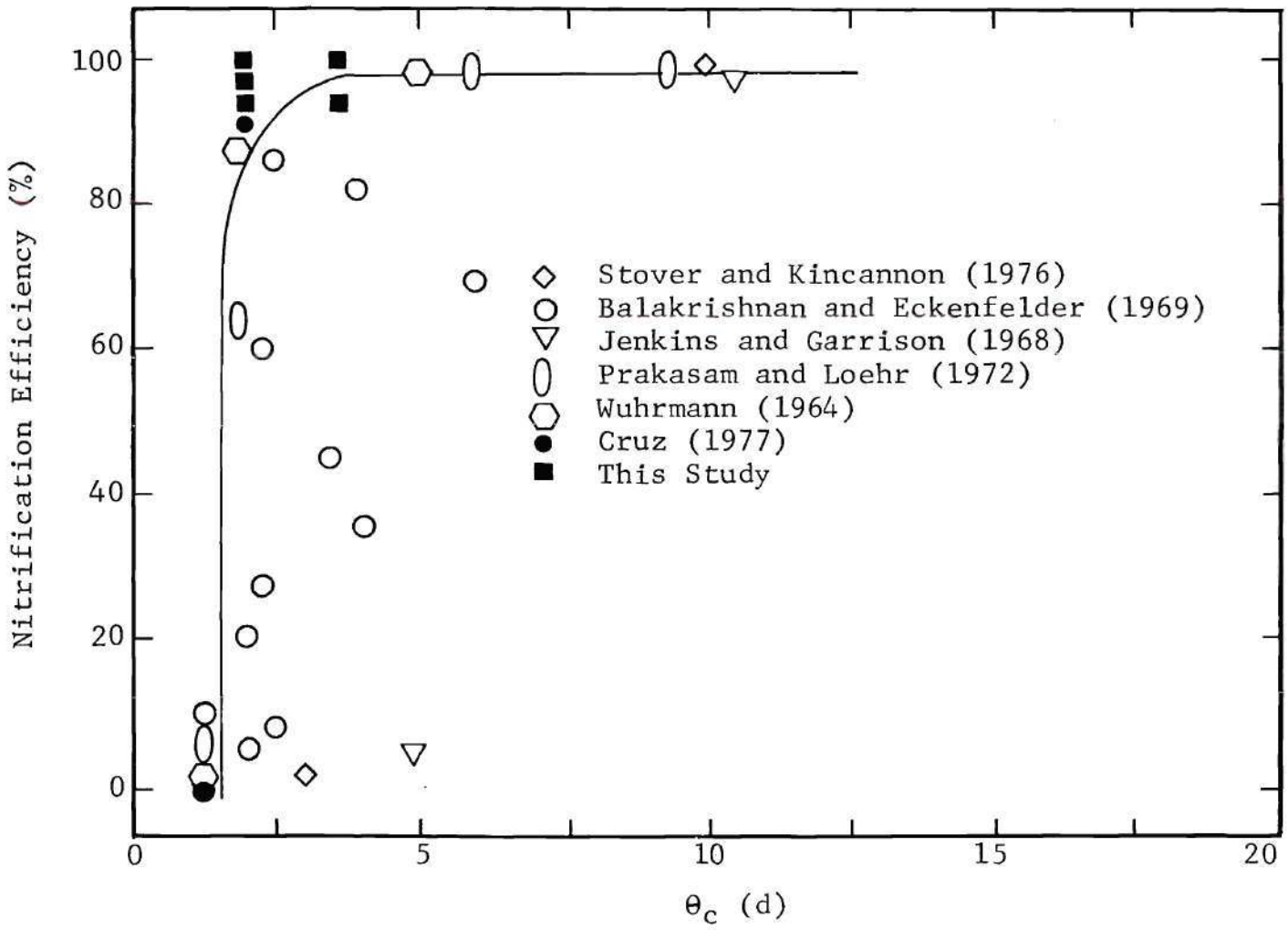

Figure 29. Comparison of Nitrification Efficiency and θ_c for Suspended- and Attached-Growth Biological Treatment Systems

this present study showed approximately the same removal of ammonia-nitrogen at similar conditions to the systems observed by Antonie (1976).

Additional performance parameters, i.e. mass flows, loading rates and removal rates, are listed in Table 14. The loading rates and the removal rates were the same in all the experiments except run 1, when nitrification did not occur. As the influent COD mass flow was increased the effluent nitrate-nitrogen mass flow generally decreased and the mass flow of the TKN in the sludge generally increased. This phenomenon resulted from more nitrogen being used in synthesis by heterotrophs as the organic load was increased.

Pretorius (1971) achieved an overall ammonia-nitrogen removal rate of 0.28 d^{-1} with influent COD and ammonia-nitrogen concentrations of 210 mg/1 and approximately 39 mg/1, respectively, while the HRT was 110 min. Phase VA had a similar ammonia-nitrogen removal rate of 0.27 d^{-1} with influent COD and ammonia-nitrogen concentrations of 172 mg/1 and 20.2 mg/1 and a HRT of 120 min. These two RBC systems achieved very similar results.

The influent and effluent nitrogen species were summed and listed in Table 15 to enable a nitrogen balance around the RBC during each experiment. The influent nitrogen was

Table 14

Additional Performance Parameters of Nitrification

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
Influent Nitrogen ($g/d \cdot m^2$)	1.11	1.62	1.24	1.44	1.41	1.30	1.17	1.45	1.41
Effluent Nitrogen ($g/d \cdot m^2$)									
NH_4-N	0.01	0.02	0.03	0.01	0.02	0.05	0.97	0.07	0.01
NO_2-N	—	0.002	—	—	—	—	0.001	0.06	—
NO_3-N	1.02	1.05	1.25	1.12	1.11	0.88	0.01	1.35	1.08
TKN[2]	0.19	0.30	0.19	0.39	0.30	0.48	0.12	0.20	—
Loading Rate (day^{-1})	0.23	0.22	0.41	0.23	0.27	0.16	0.74	0.26	—
Removal Rate (day^{-1})	0.23	0.22	0.41	0.23	0.27	0.16	0.13	0.25	—

LEGEND: 1 - Data from Cruz (1977)

2 - TKN in cells leaving the system (effluent, scraping and sieving)

— Data not collected

Table 15

Nitrogen Balance

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
Influent Nitrogen ($g/d \cdot m^2$)	1.11	1.62	1.24	1.44	1.41	1.30	1.17	1.45	1.41
Effluent Nitrogen ($g/d \cdot m^2$)	1.22	1.37	1.47	1.52	1.43	1.41	1.10	1.68	1.36[2]
Difference[3] (%)	+10	-15	+19	+6	+1	+8	-6	+16	-4

LEGEND: 1 - Data from Cruz (1977)

2 - TKN was not measured; therefore, it was assumed to be 0.27 $g/d \cdot m^2$ (the average for the other eight experiments) in the cells that left the system.

3 - + = increase in nitrogen across the system
- = decrease in nitrogen across the system

approximately the same as the effluent nitrogen during phases I, IV, VA and VB, and runs 1 and 3, i.e. the balance was within 10%. The effluent nitrogen was considerably higher than the influent nitrogen in phase III and run 2; and lower in phase II. The most probable reason the data show an increase in nitrogen across the RBC system was inaccuracy in the analytical techniques. A decrease in nitrogen concentration from 1.62 $g/d \cdot m^2$ to 1.37 $g/d \cdot m^2$ in phase II may be due to denitrification in an anerobic layer; however, this phenomenon was only observed in one phase. From the data presented in Table 15, it was concluded that denitrification was not a significant biological process occurring in the biofilm.

Sludge Production

Another significant area of importance in evaluating a biological system is sludge production. A summary of the sludge data is presented in Table 16. The MLSS in the RBC were kept at low concentrations by the circulation system. The MLSS during the experiments ranged between 33 mg/l and 487 mg/l. These low MLSS are representative of full-scale RBC systems. Antonie (1976) reported MLSS concentrations in full-scale RBC systems between 49 mg/l and 275 mg/l.

The sludge production generally increased as the

Table 16

Summary of Sludge Data

	PHASE						RUN[1]		
	I	II	III	IV	V A	V B	1	2	3
θ_c (day)	3.5	3.5	2.0	2.0	2.0	2.0	1.3	2.0	—
Sludge Production ($g/d \cdot m^2$)	1.40	1.97	1.61	3.09	2.58	3.92	1.23	2.33	—
Yield (gm of sludge produced per g of COD removed)	0.37	0.39	0.29	0.36	0.23	0.31	0.40	0.53	—
Nitrogen in Sludge (%)	13.5	12.3	11.9	12.7	11.8	12.2	9.7	8.7	—
MLSS (mg/l)	242	290	198	363	330	487	33	212	—
Biomass per Area (g/m^2)	3.22	5.23	1.80	3.95	3.19	4.89	1.39	4.24	—

LEGEND: 1 - Data from Cruz (1977)

— Data not collected

influent COD mass flow increased (see Figure 30). The yield generally decreased with increased influent COD, as shown in Figure 31. The yield observed during this present study ranged between 0.23 and 0.53 g of sludge produced per g of COD removed. Antonie (1976) reported similar yields that ranged between 0.24 and 0.75 g of sludge produced per g of BOD removed with an RBC system achieving 90 to 95 percent BOD removal.

The sludge production and yield were not observed to be dependent on θ_c. However, the percent nitrogen in the sludge generally increased with θ_c (see Figure 32). Some percent nitrogen values reported by other researchers are listed in Table 17. The results observed in this present study correspond to those reported by Hoover and Porges (1952) and Eckenfelder and O'Connor (1961).

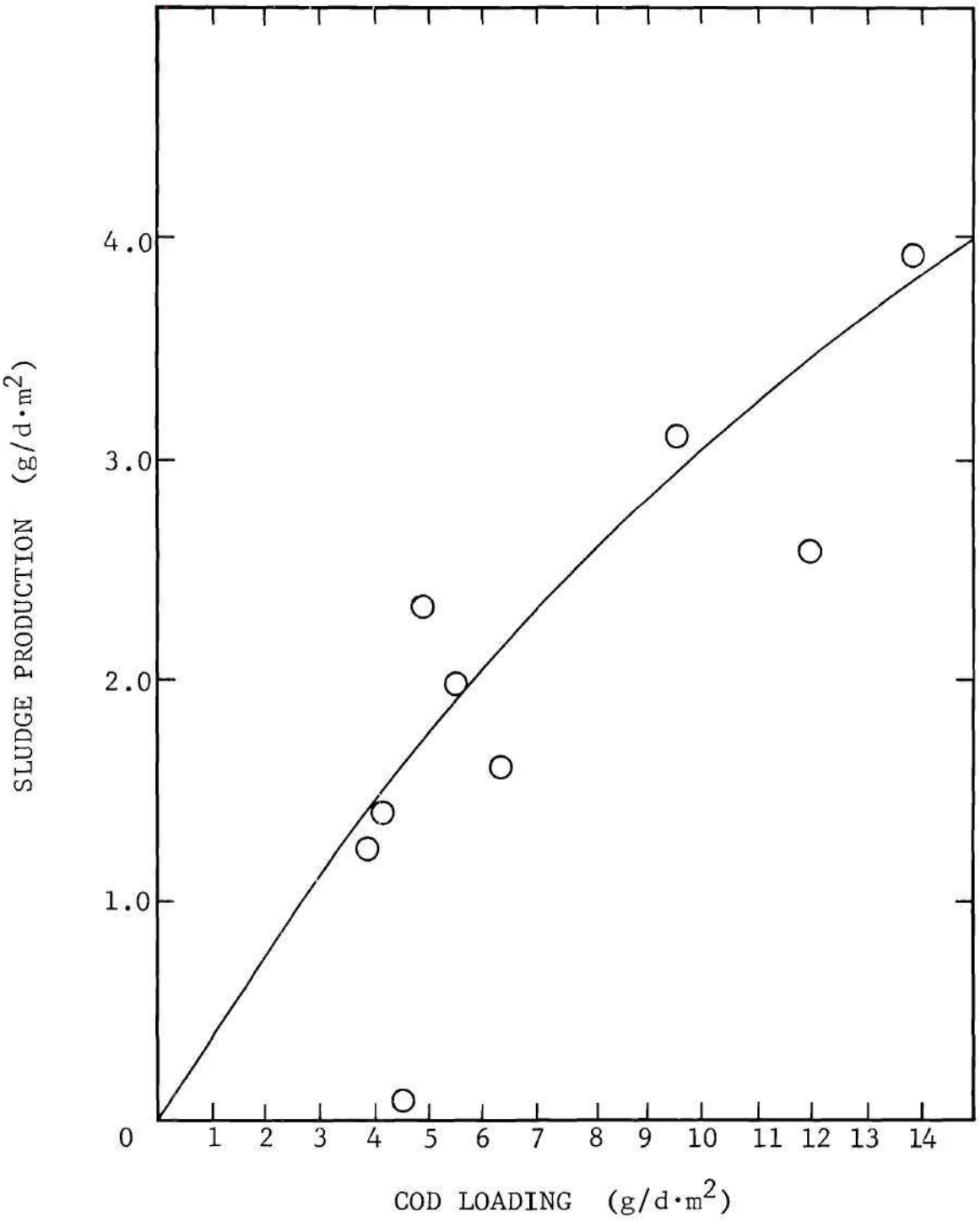

Figure 30. COD Loading Vs. Sludge Production

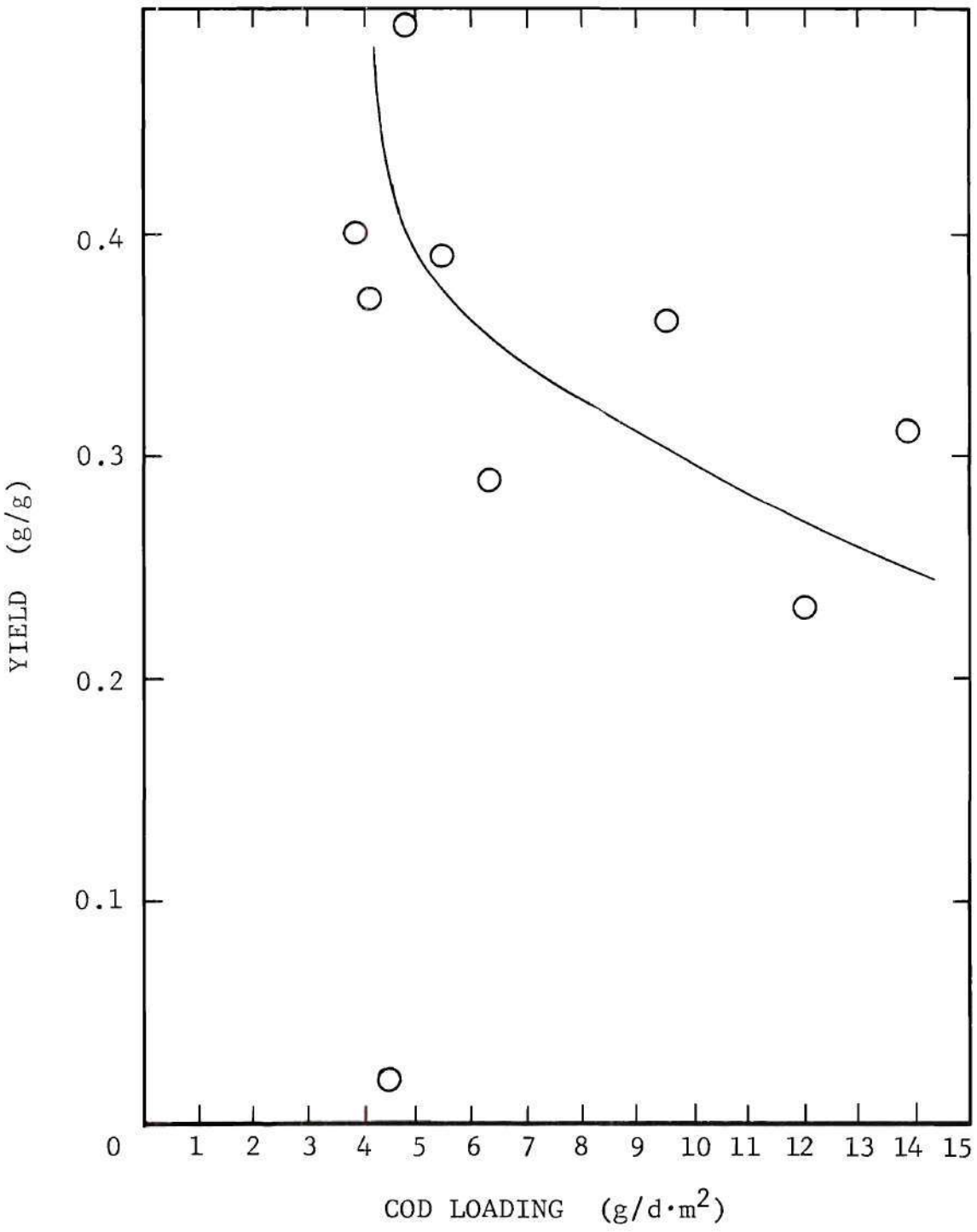

Figure 31. COD Loading Vs. Yield

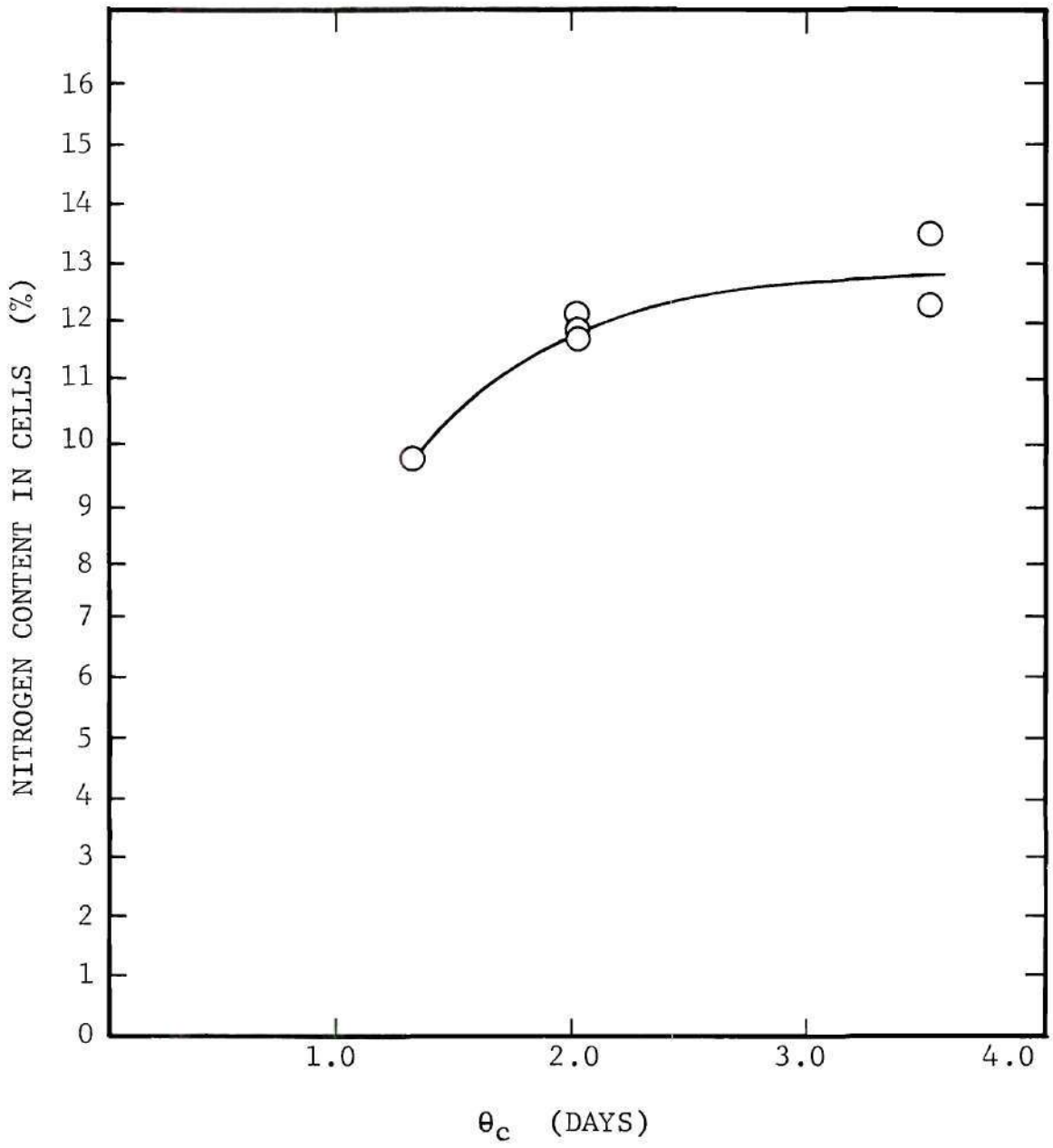

Figure 32. θ_c Vs. Nitrogen Content in Cells

TABLE 17

REPORTED NITROGEN CONTENT IN DIFFERENT SLUDGES

Type of Sludge	% N	Source
Raw Primary	2.4 - 2.9	Vesilind (1974)[1]
Trickling Filter	2.9	Vesilind (1974)[1]
Activated Sludge	3 - 5.6	Vesilind (1974)[1]
Skim Milk Activated Sludge	11.27	Hoover & Porges (1952)
Biological Sludges	8 - 15	Eckenfelder & O'Conner (1961)
Yeast	8.5	Eckenfelder & O'Conner (1961)
Activated Sludge	8.0	Helmers et al. (1952)
Domestic and Digested Mixed Sludge	1.8 - 5.9	Vesilind (1974)[1]

[1]As reported by Vesilind (1974) from other sources.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

Conclusions

The following conclusions were based on the results observed from the laboratory bench-scale RBC system used in this study.

(1) Organic loadings as high as 13.7 g COD/d·m^2 were observed to have no effect on nitrification with an influent ammonia-nitrogen concentration and hydraulic loading of 1.3 g N/d·m^2 and 72 l/d·m^2, respectively.

(2) The hydrolysis of organically-bound nitrogen (as glycine) did not limit nitrification.

(3) The critical θ_c value for washout of nitrifying microorganisms was observed to be approximately 1.5 d. These results agree with data reported on pure cultures of Nitrosomonas and Nitrobacter and mixed cultures in wastewater treatment systems.

(4) The θ_c concept was effective in characterizing the microbial population in this RBC system and is a

potential method for modeling RBC systems.

Recommendations

To better understand and use the RBC system for wastewater treatment, the following actions are recommended:

(1) Biofilms need further investigation with respect to growth rates, sloughing and active/inactive layers.

(2) The RBC system operated in this study was never observed to be oxygen limited. The system should be operated at higher organic and nitrogen loading to determine the loading limitations of the system.

(3) To model the RBC process, further investigation is required as to the effects of dissolved oxygen, disc rotational speed and temperature.

APPENDIX

Table A-1

Summary of the Total Suspended Solids Fractions, Phase I

DAY	SS_D (grams)	SS_C (grams)	SS_M (grams)	SS_E (grams)	SS (grams)
101					
102					
103					
104					
105					
106					
107					
108	0.84	1.32	0.93	0.03	3.12
109					
110					
111	1.08	1.14	0.84	0.03	3.09
112					
113					
114	0.90	1.32	1.05	0.02	3.29
115					
116					
117	0.96	1.20	1.32	0.06	3.54
118					
119					
120	0.78	0.96	0.54	0.04	2.32
121					
122					
123	0.72	0.57	0.72	0.03	2.04
124					
125					
126	1.35	0.87	1.23	0.06	3.51
127					
128					
129	1.14	1.59	0.84	0.06	3.63
130					
131					
132	1.26	2.12	1.02	0.21	4.61
133					
134					
135	1.23	1.85	0.99	0.46	4.53
136					
137					
138	1.17	2.52	1.38	0.59	5.66
139					
140					
141	0.87	2.31	0.60	0.33	4.11
142					
143					
144	0.63	2.02	0.66	0.33	3.64
145					
146					
147	1.20	2.16	1.08	0.44	4.88
148					

Table A-2

Summary of the Total Suspended Solids Fractions, Phase II

DAY	SS_D (grams)	SS_C (grams)	SS_M (grams)	SS_E (grams)	SS (grams)
149					
150	0.84	1.11	0.60	0.54	3.09
151					
152					
153	1.05	0.92	0.54	0.59	3.10
154					
155					
156	0.93	2.16	1.44	0.77	5.30
157					
158					
159	1.50	1.47	1.08	0.86	4.91
160					
161					
162	1.14	1.39	1.02	0.44	3.99
163					
164					
165	1.08	1.64	1.68	0.85	5.25
166					
167					
168	1.20	1.07	0.39	0.24	2.90
169					
170					
171	2.13	1.48	0.81	0.48	4.90
172					
173					
174	2.16	1.45	1.59	1.10	6.30
175					
176					
177	1.32	2.64	1.80	1.06	6.82
178					
179					
180	1.89	2.07	1.50	1.17	6.63
181					
182					
183	1.77	1.80	0.87	1.06	5.50
184					
185					
186	2.28	2.57	1.05	1.84	7.74
187					
188					
189	1.89	2.71	0.72	0.61	5.93
190					
191					
192	1.92	4.14	1.56	0.42	8.04
193					
194					
195	1.83	1.63	0.57	0.40	4.43
196					
197					
198	1.56	3.22	1.11	0.73	6.62
199					
200					
201	1.83	4.48	1.65	0.72	8.68
202					
203					
204	1.74	1.80	0.69	0.62	4.85
205					
206					
207	2.10	2.11	1.26	0.45	5.92

Table A-3. Summary of the Total Suspended Solids Fractions, Phase III

DAY	SS_D (grams)	SS_C (grams)	SS_M (grams)	SS_E (grams)	SS (grams)
208					
209	2.35	1.14	0.84	0.23	4.56
210					
211	2.22	0.45	1.16	0.67	4.50
212					
213	2.86	0.67	1.78	0.18	5.49
214					
215	2.76	2.23	0.96	0.22	6.17
216					
217	0.90	1.80	1.06	0.37	4.13
218					
219	0.80	1.29	0.78	0.27	3.14
220					
221	0.68	1.61	0.78	0.93	4.00
222					
223	0.94	2.59	1.06	0.60	5.19
224					
225	0.86	2.66	0.68	0.46	4.66
226					
227	0.86	2.81	0.72	0.27	4.66
228					
229	0.72	1.92	0.80	0.39	3.83
230					
231	0.78	2.34	0.80	0.39	4.31
232					
233	0.66	1.90	0.52	0.26	3.34
234					
235	0.50	1.95	0.74	0.27	3.46
236					
237	0.60	0.90	0.54	0.52	2.56
238					
239	0.64	1.74	0.72	0.28	3.38
240					
241	0.52	1.92	0.62	0.28	3.34
242					
243	0.82	1.84	0.96	0.22	3.84

Table A-4. Summary of the Total Suspended Solids Fractions, Phase IV

DAY	SS_D (grams)	SS_C (grams)	SS_M (grams)	SS_E (grams)	SS (grams)
244					
245	1.06	1.87	0.44	0.28	3.65
246					
247	0.88	1.81	0.88	0.32	3.89
248					
249	1.30	3.25	1.34	0.44	6.33
250					
251	1.52	2.77	1.40	0.47	6.16
252					
253	1.08	2.75	0.98	0.58	5.39
254					
255	1.52	1.90	3.08	1.08	7.58
256					

Table A-5. Summary of the Total Suspended Solids Fractions, Phase V

DAY	SS_D (grams)	SS_C (grams)	SS_M (grams)	SS_E (grams)	SS (grams)
257	1.20	2.52	0.82	0.57	5.11
258					
259	0.90	2.66	0.96	0.19	4.71
260					
261	1.06	2.43	1.44	0.24	5.17
262					
263	0.94	3.36	1.54	0.14	5.98
264					
265	1.20	2.65	1.54	0.13	5.52
266					
267	1.02	1.83	2.12	0.28	5.25
268					
269	1.36	2.94	0.90	0.33	5.53
270					
271	1.24	3.35	1.10	0.34	6.03
272					
273	2.10	2.80	1.36	0.47	6.73
274					
275	1.36	5.63	2.60	0.37	9.96
276					
277	1.34	6.41	1.56	1.12	10.43
278					
279	2.36	1.26	3.16	0.46	7.24
280					

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