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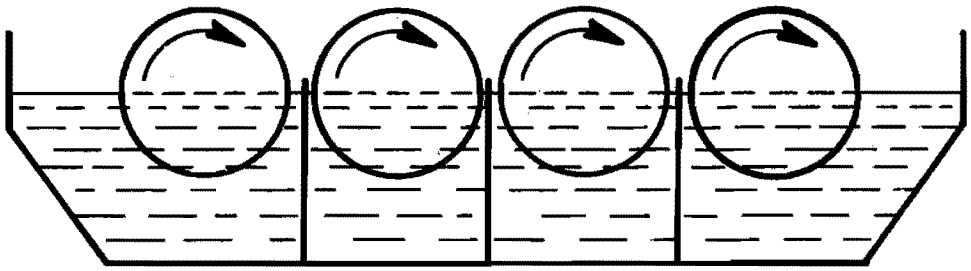

Utah Water Research Laboratory
College of Engineering
Utah State University
Logan, Utah 84322

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THE KINETICS OF ROTATING BIOLOGICAL CONTACTORS TREATING DOMESTIC WASTEWATER

by

Abraham Pano
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ABSTRACT

Four, four-stage, laboratory scale rotating biological contactor (RBC) units were employed to develop kinetic models for the RBC process when treating domestic wastewater. The study was conducted at three different temperatures of 5°C, 15°C, and 20°C. Each unit operated at a different constant organic loading rate that ranged from 4 gCOD/m^2/d to 14 gCOD/m^2/d and at a constant detention time of 2 hours. Carbonaceous substrate removals measured as COD concentrations with the four-stage RBC's were 80, 85, and 90 percent for 5°C, 15°C, and 20°C, respectively. The major part of the removal occurred in the first stages. The overall percentage removals of ammonia nitrogen were 87 to 98 percent at 15°C and from 91 to 99 percent at 20°C depending upon the influent organic loading rate. At 5°C there was no ammonia nitrogen removal.

Kinetic models were developed and kinetic constants were determined for COD and ammonia nitrogen removal in the first and succeeding stages of the RBC units. Biomass yield, biomass stabilization, and ammonia nitrogen removal were also evaluated.

Monod growth kinetics were used in the development of the models for carbonaceous substrate removal in the first stages and for ammonia nitrogen removal in the system. The temperature dependency of each kinetic constant was determined for the range of 5°C to 20°C.

Steady-state kinetic models were developed, and kinetic constants were determined as a function of temperature to provide a rational design approach for the RBC process treating domestic wastewater.

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INTRODUCTION

Nature of the Problem

The rotating biological contactor (RBC) has been employed for biological treatment of municipal wastewater for several years. The process has been shown to be efficient in carbonaceous substrate removal and ammonia nitrogen removal. In recent years, the use of the RBC process has increased mainly because of the simplicity of operation and the low power consumption.

Traditionally, the design of RBC systems has been based primarily on empirical relationships between pollutant removal efficiency and hydraulic loading rates per unit surface area of the RBC discs. Presently, the design hydraulic loading rates are adjusted by a safety factor for wastewater temperatures below 12.8°C (55°F).

This empirical design approach ignors the basic concepts of biological substrate removal kinetics. Although several recent investigations have developed models to describe the performance of the RBC process, there are limited data available for setting values for the kinetic constants associated with substrate removal. Even information concerning the effect of temperature on the kinetic constants associated with RBC substrate removal is scarce.

Objectives

The general objective of this study was to develop a mathematical model for carbonaceous and ammonia nitrogen substrate removal as a function of temperature in an RBC system treating domestic wastewater.

The specific objectives were:

1. To develop kinetic models of the degrees of treatment achieved by the different processes associated with carbonaceous and ammonia nitrogen removal in the first and following stages of an RBC system.

2. To determine the kinetic constants for each process at each stage and temperature.

3. To determine the effect of temperature on the kinetic constants.

LITERATURE REVIEW

General Background

The rotating biological contactor (RBC) process is a secondary biological wastewater treatment system, designed to treat municipal and industrial wastewaters. The raw wastewater first passes through a primary clarifier. The secondary treatment utilizes biological growth attached to media on closely spaced rotating contactor disks mounted on shafts and submerged approximately 40 percent in the wastewater. As the disks rotate, the attached biomass is alternately exposed to wastewater in the tank and to air above, providing simultaneous oxygen supply and substrate extraction from the wastewater stream for growth and maintenance of the attached organisms. When the growth becomes large enough, fluid shear forces cause portions of biomass to be stripped from the contactor media into the wastewater stream. The sloughed biomass is then separated from the treated carrier stream in a secondary clarifier. The carrier stream effluent is discharged or further treated, and the biological sludge is treated with appropriate processes.

Full scale RBC installations generally consist of four to six stages in series, each stage with several disks rotating on a shaft (Figure 1). The contactor media on the disks are made of plastic material such as high density, polyethylene, closely packed, and mounted on a shaft. The contactors range from 1.2 to 3.6 m (4-12 ft) in diameter and 1.5 to 6 m (5-20 ft) in length. The contactors rotate at two to four revolutions per minute (rpm) with partial submergence (about 40 percent) in a semicircular or rectangular tank made of concrete or steel. RBC plants include primary treatment facilities and secondary clarifiers.

Applications of the RBC process for secondary treatment is relatively new. Although the historical background of the process starts at the beginning of the twentieth century (Grieves 1972; Antonie 1976; Smith and Bandy 1980), the first commercial RBC installation in the USA went into operation in 1963 (Antonie 1976).

The RBC process is considered relatively stable and provides a high degree of carbonaceous substrate removal efficiency (90 percent $\pm$ 5 percent BOD_5 removal) and a substantial degree of nitrification (Antonie 1976). The modular construction of the RBC provides flexibility in the plant configuration and in upgrading existing plants. Finally, the RBC is simple to operate, and energy requirements are much less than the energy requirements in activated sludge systems: 0.212 kwh/1000 m^3 in RBC compared to 0.708 kwh/1000 m^3 in activated sludge (Chesner and Iannone 1980).

In 1979 there were 263 municipal and 58 industrial RBC installations in the USA (Chesner and Iannone 1980). Most of these units were treating a wastewater flow rate of less than 18,900 m^3/d (5 mgd).

Process Performance

Carbonaceous substrate removal

Municipal wastewater treatment. During the last 10 years, the RBC process has been investigated intensively. Bench scale units, pilot plant units, and full-scale plants have been evaluated for carbonaceous substrate removal and for nitrification. The first studies in the USA of the RBC process

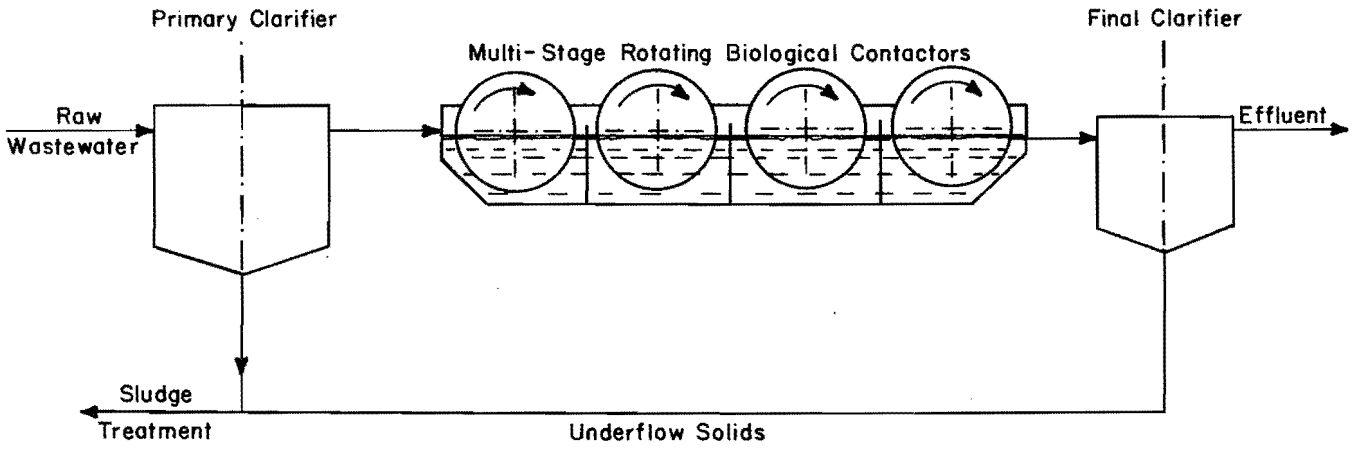


Figure 1. Flow diagram for rotating contactor process.

treating municipal wastewater were conducted during 1969-1971. Two-stage and four-stage, 1.75 m (5.74 ft) diameter RBC units with secondary clarifiers were studied to determine the effects of hydraulic loading rate, rotational speed, and the number of stages (Antonie 1976). The dependence of the 5-day biological oxygen demand (BOD_5) removal on the hydraulic loading rate which could be effectively treated was mathematically defined by a first order kinetic relationship. Some decline in the kinetic constant was observed beyond 90 percent BOD_5 removal. Most of the removal occurred in the first two stages. An increase from two to four stages resulted in a 5 to 10 percent increase in BOD_5 removal.

The study concluded that the optimum rotational speed for BOD_5 and ammonia nitrogen removal was 3.2 rpm, i.e., about 18 m/min (59 ft/min) peripheral velocity. It should be emphasized, however, that at high loading rates [0.2 $m^3/m^2/d$ (5 gpd/sq ft) was tested] higher rotational speeds increased the BOD_5 removal about 10 percent. The maximum performance reported by Antonie (1976) was about 95 percent BOD_5 removal (with four stages) at 0.04 $m^3/m^2/d$ (1 gpd/sq ft), and removals declined to 75 percent at approximately 0.2 $m^3/m^2/d$ (5 gpd/sq ft).

A 0.6 m diameter (2 ft) pilot plant operating with volume to surface area ratios in the range of 2.7 l/m^2 to 13 l/m^2 (0.067 gal/ft^2-0.32 gal/ft^2) showed better BOD_5 and ammonia nitrogen removals as the volume to surface area ratio was increased up to a value of 4.9 l/m^2 (0.12 gal/ft^2) (Antonie 1976). Sludge production was within the range of 0.5-0.6 grams/gram of BOD_5 removed. It was observed also that there was a trend toward an increase in this ratio as the temperature and the BOD_5 removal decreased.

Torpey et al. (1971) treated primary municipal wastewater effluent with a 10-stage, 0.9 m (3 ft) diameter RBC with an overall detention time of 50-60 minutes, and achieved 93 percent BOD_5 removal, 79 percent chemical oxygen demand (COD) removal, and 92 percent suspended solids (SS) removal. The staged operation showed plug flow characteristics, first order kinetics for the carbonaceous substrate removal, and a decline in the kinetic parameters beyond the fifth stage. During the experiments, the attached biomass thickness was maintained below 1.5 mm (0.059 in) by periodic removal of the biomass using a water jet. Although the experiments were conducted at a rotational speed of 10 rpm, or a corresponding peripheral velocity of 28 m/min (91.8 ft/min), dissolved oxygen concentrations in the wastewater in the first two stages were less than 1 mg/l.

Borchardt (1971) used a three-stage, 1.2 m (4 ft) diameter RBC, with an overall detention time of 21 minutes, to treat municipal wastewater. Overall BOD_5 removals of 83 percent to 94 percent (including primary treatment) were achieved. Malhotra et al. (1975) showed that at 13°C (56°F) and a hydraulic loading rate of 0.11 $m^3/m^2/d$ (2.65 gpd/sq ft) the influent BOD_5 concentration of 85 mg/l applied to the first stage was reduced to 13 mg/l at the third stage, i.e., 84.7 percent removal (without secondary clarification). The removal decreased to 68 percent at the same temperature with a hydraulic loading rate of 0.16 $m^3/m^2/d$ (4 gpd/sq ft), and an influent BOD_5 concentration of 100 mg/l. In both cases the major removal occurred in the first stage (40-50 percent).

Pescod and Nair (1972) incorporating an anaerobic section beneath the RBC investigated the feasibility of the process for tropical countries. The bench scale unit consisted of a five-stage RBC with 19 cm (7.5 in) diameter perforated PVC discs. The mean theoretical hydraulic detention time could be varied from 2 to 8 hours. The domestic wastes that were treated with the system had a COD concentration of 400 mg/l, and more than 80 percent COD removal was attained within 2 hours. The system performed as a plug flow unit with first order kinetics.

Pretorius (1971) introduced the concept of carbonaceous substrate removal per unit biomass. A nine-stage, 40 cm (15.7 in) diameter bench scale unit treating anaerobic digester effluent at 20°C was evaluated. Approximately 44 percent of the total maximum COD concentration of 220 mg/l was reduced in the first stage. The results indicated that 0.49 g of COD was removed per gram of attached biomass per day or 12.7 g of COD per m^2 of media surface area per day. The maximum attached biomass was reported to be 45 g/m^2.

The first full-scale use of an RBC system to treat domestic wastewater in the USA was located at Pewaukee, Wisconsin (Antonie et al. 1974). The flow rate was 1900 m^3/day (0.5 mgd). The four-stage, 3.1 m (10 ft) diameter RBC followed by a secondary clarifier achieved 85 percent BOD_5 removal at a hydraulic loading rate of 0.11 $m^3/m^2/d$ (2.8 gpd/sq ft). The system treated primary effluent with a BOD_5 concentration of 100-170 mg/l. BOD_5 removals were 95 percent at a hydraulic loading rate of 0.02 $m^3/m^2/d$ (0.49 gpd/sq ft). When treating primary influent, BOD_5 concentrations of 75-98 mg/l, removals decreased about 5 percent. Significant changes in mixed liquor suspended solids (MLSS) did not occur from stage to stage.

Malhotra et al. (1975) studied the performance of a six-stage, full scale, 3800 m^3/d (1 mgd) RBC plant, treating municipal wastewater at Gladstone, Michigan. The plant operated at a mean temperature of 12.8°C (55°F), and the hydraulic loading rate was 0.06 $m^3/m^2/d$ (1.52 gpd/sq ft). The mean BOD_5 concentration of the primary effluent treated by the RBC was 114 mg/l. Utilizing final chemical precipitation with alum and polymer in the secondary clarifier, a mean BOD_5 removal of 86.9 percent was obtained. An analysis of the performance between stages showed that the MLSS did not change from stage to stage, but the same reduction in filtered BOD_5 and in particulate BOD_5 occurred from stage to stage. The reduction in the ratio of particulate BOD_5 to SS through the stages indicated stabilization of the sloughed biomass as the material passed through the stages.

An analysis of stage data (BOD-filtered, COD-filtered, and BOD-particulate) indicated first order kinetics with a decline in the kinetic parameters beyond stage three.

A different mode of operation was investigated by Ellis and Banaga (1976). A bench scale, seven-stage, 20 cm (7.9 in) diameter RBC unit was operated so that the sloughed biomass could be periodically wasted from the bottom of the tank. Settled domestic wastewater with a mean BOD_5 concentration of 240 mg/l was applied to the system. A maximum BOD_5 removal of 97.5 percent was observed when the RBC was operating with a 3 hr detention time, 26 percent media submergence

and a rotational speed of 4 rpm. The system operated efficiently when loaded in excess of 20 $gBOD_5/m^2/d$.

Almost the same mode of operation used by Ellis and Banaga (1976) was investigated by Gutierrez et al. (1980). A four-stage, 3.6 m (12 ft) diameter RBC was used to upgrade primary effluent at Edgewater, New Jersey. The primary settling tank was converted to an RBC underflow clarifier by installing an intermediate floor. Primary effluent consisting of municipal wastewater combined with storm water was treated in the RBC underflow clarifier unit. With influent BOD_5 concentrations of 140 mg/l, 85 percent removal was observed and the optimum loading was found to be 9-11 $gBOD_5/m^2/d$.

Dupont and McKinney (1980) analyzed the performance data from a full scale, 19,000 m^3/d (5 mgd), five-stage, 3.3 m (11 ft) diameter RBC plant, treating municipal wastewater at Kirksville, Missouri. The overall BOD_5 removal was approximately 90 percent at a hydraulic loading rate of 0.05 $m^3/m^2/d$ (1.3 gpd/sq ft) and an influent BOD_5 of 192 mg/l. At the same hydraulic loading rate and higher influent BOD_5 concentrations, BOD_5 removal increased to 94 percent. During one period of operation, the influent to the RBC contained a highly soluble waste that was thought to come from a creamery in the city. The BOD_5 removal during this period was reduced to 81 percent but the influent BOD_5 concentration was 383 mg/l. BOD_5 removal per unit area was correlated with the influent organic load per unit area of media, and the slope of the relationship was 0.893 and was significant at the 5 percent level.

Given (1980) used a five-stage, 1.09 m diameter RBC unit to treat a dilute primary wastewater (influent BOD_5 of 11 to 34 mg/l), and achieved a BOD_5 removal (including secondary clarification) of 92 percent at a hydraulic loading rate of 0.056 $m^3/m^2/d$. BOD_5 removal was decreased to 50 percent at a hydraulic loading rate of 0.248 $m^3/m^2/d$.

Industrial wastewater treatment. Banerji (1980) analyzed BOD_5 removal data obtained with a four-stage RBC unit treating various types of industrial wastes and found a linear declining relationship between BOD_5 removal and hydraulic loading rate. Similar results have been cited for RBC systems treating domestic wastewater (Antonie 1976). In contrast to the results obtained with RBC systems treating domestic wastewater, the percent BOD_5 removal decreased as the influent BOD_5 concentration increased.

The first industries to apply the RBC process to wastewater treatment were food processing industries (dairy, bakery, winery, meat slaughtering, etc.) and other organic processing industries such as paper mills and oil refineries. Welch (1968) using synthetic dairy waste with a two-stage RBC having a 30-minute detention time in each unit, incorporated interstage clarification and achieved more than 60 percent COD removal with influent COD concentrations of 500 mg/l. COD removal percentages decreased with increasing influent concentration and decreasing detention time.

The effect of loading rate and rotating speed was demonstrated by Chittenden and Wells (1971), using a three-stage, 1.2 m (4 ft) diameter RBC pilot unit, treating anaerobic effluent. With a rotating speed of 3 rpm and a hydraulic loading rate of 0.33 $m^3/m^2/d$ (8 gpd/sq ft), the overall BOD_5 removal after the third stage was 49.7 percent and BOD_5 removal in the first stage was 42.7 percent. When the rotating speed in the first stage was increased to 6 rpm, the percentage BOD_5 removal increased to 64.5 percent with 52.5 percent removal occurring in the first stage. The overall BOD_5 removal was increased to 83.2 percent (79.5 percent in the first stage) at a hydraulic loading rate of 0.17 $m^3/m^2/d$ (4 gpd/sq ft).

Antonie and Hynek (1973) presented a summary of the percent BOD_5 removal achieved with an RBC pilot plant treating different processing wastes. BOD_5 removals were reported to be 71 percent to 98 percent.

Mikula (1979) used a four-stage, 1.2 m (4 ft) diameter RBC pilot plant to treat cheese-processing wastewater, and the process removed 88 percent of the influent COD at hydraulic loading rates of 0.012 $m^3/m^2/d$ (0.3 gpd/sq ft). At a hydraulic loading rate of 0.028 $m^3/m^2/d$ (0.69 gpd/sq ft), 71.1 percent of the influent COD was removed.

Ammonia nitrogen removal

Results from pilot plant and full scale RBC units have demonstrated that there is substantial nitrification in the RBC process when operated as a secondary treatment process. Antonie (1976) and Antonie et al. (1974), reporting the nitrification performance with pilot and full scale RBC units at Pewaukee, emphasized the following performance characteristics:

- As the mixed liquor BOD_5 decreases below 40 mg/l, the nitrification process is stimulated.
- Ammonia nitrogen removal obeys first order kinetics up to at least 97 percent removal.
- Ammonia nitrogen removal is affected by the rotational speed and volume to surface area ratio in the same way that carbonaceous substrate removal is affected.
- Full scale plant results indicate at hydraulic loading rates of 0.074 $m^3/m^2/d$ (1.8 gpd/sq ft) a 25-45 percent ammonia nitrogen removal was achieved.

The low ammonia nitrogen removal values were obtained at influent BOD_5 concentrations of 100-170 mg/l. Ammonia nitrogen removal was approximately 95 percent at hydraulic loading rates of 0.025 $m^3/m^2/d$ (0.6 gpd/sq ft). Torpey et al. (1971) achieved only 56 percent removal with 10 stages even though in the final stages BOD_5 concentrations were less than 9 mg/l. The relatively low performance obtained by Torpey et al. (1971) may be attributable to the special operation of biomass scraping.

A six-stage, 59 cm (23.2 in) diameter RBC unit, treating a synthetic substrate at 20°C, achieved 100 percent ammonia nitrogen removal with an influent COD concentration of 250 mg/l and an ammonia nitrogen concentration of 27.6 mg/l (Stover and Kincannon 1975). The system was hydraulically loaded at 0.02 $m^3/m^2/d$ (0.5 gpd/ft^2). Murphy et al. (1977) used a 0.5 m (1.64 ft) diameter RBC unit operated at 13 rpm to treat screened municipal wastewater, and achieved a nitrification rate of approximately 45 mg total Kjeldahl nitrogen (TKN)/m^2/hr.

Hitdlebaugh and Miller (1980) described the nitrification performance of a full scale [22,700 m^3/d (6 mgd)] domestic wastewater RBC plant. The plant had six stages and was operated at a hydraulic loading rate of 0.033 $m^3/m^2/d$ (0.82 gpd/sq ft) at a temperature of 26°C and at a hydraulic loading rate of 0.04 $m^3/m^2/d$ (1 gpd/sq ft) at 13°C. The plant achieved better ammonia nitrogen removal at 13°C (75.6 percent ammonia removal through the six stages of the RBC) than at 26°C (58.8 percent ammonia removal), with the same concentrations of ammonia nitrogen in the influent. The low concentrations of dissolved oxygen (DO) in the mixed liquor during the summer (0.7-2.2 mg/l) contributed to the poorer warm temperature performance. During the winter the DO concentrations were 2.7-4.8 mg/l.

The full scale plant at Gladstone, Michigan (Hynek and Iemura 1980), was reported to achieve 93 percent ammonia nitrogen removal during the summer (17°C) and 67.3 percent removal during the winter (9°C). Ito and Matsuo (1980) worked with 25 cm (9.8 in) diameter RBC units treating synthetic substrate and reported that nitrification could not proceed when the COD loading exceeded 20 $gCOD/m^2/d$.

Using a four-stage, 0.6 m (2 ft) diameter RBC unit to treat primary municipal wastewater, Zenz et al (1980) observed that ammonia nitrogen removal efficiency was directly correlated with hydraulic detention time (1 to 5 hrs) and temperature (9°C to 25°C) and inversely related to the ammonia nitrogen loading rate (0.24 $g/m^2/d$-2.28/m^2/d). Ammonia nitrogen removals greater than 90 percent were achieved at hydraulic detention times of 4 hrs or greater, at ammonia nitrogen loading rates less than 0.54 $g/m^2/d$ (0.11 $lb/1000\ ft^2/d$), and only at temperatures higher than 18°C.

Temperature effects on RBC performance

Based upon experience with other types of biological wastewater treatment units, the efficiency of RBC units would be expected to increase at higher temperatures. However, the decrease in dissolved oxygen saturation concentrations at higher temperatures results in a lower aeration capacity; and for the RBC system, this effect may offset the increase in biological activity due to a temperature increase.

The percent BOD_5 removals observed at the full scale RBC system at Pewaukee, Wisconsin, declined at a temperature below 12.7°C (55°F) (Antonie et al. 1974). Between 12.7°C to 7°C (45°F) BOD_5 removals declined 3-7 percent depending upon the hydraulic loading rates. In a later study, Antonie (1976) stated that temperature variation between 12.7°C (55°F) and 29.4°C (85°F) did not affect BOD_5 and ammonia nitrogen removal performance, but both performances decreased below 12.7°C.

Malhotra et al. (1975) reported that the pilot plant studies conducted by Borchardt (1971) at temperatures of 8.9°C (48°F) to 17.2°C (63°F) showed approximately a 1 percent decline in BOD_5 removal for each 1°C decline in temperature. Ellis and Banaga (1976) found that with a bench scale RBC unit that an increase in temperature from 11°C to 18°C resulted in an increase of 3 percent in BOD_5 removal efficiency, while an increase in temperature from 18°C to 27°C resulted only in an increase of about 2 percent.

Murphy et al. (1977), feeding screened sewage to a 0.5 m (1.64 ft) diameter RBC unit operating in a temperature range of 7°C to 25°C, reported that the nitrification rate increased with an increase in the temperature from 15 $mg/m^2/hr$ at 7°C to about 65 $mg/m^2/hr$ at 25°C.

Wu et al. (1980) used regression analyses of available RBC data to show that the function of influent soluble BOD_5 remaining in the effluent is proportional to $T^{-0.2477}$, where T is the temperature in °C. The mixed liquor temperature range evaluated was 2°C to 13°C. When operating as a secondary treatment process, the function of influent ammonia nitrogen remaining in the effluent was proportional to $T^{-0.6229}$, where T is the temperature of the mixed liquor in °C.

The effect of low dissolved oxygen saturation values at high temperatures was demonstrated at a full scale, six-stage RBC plant (Hitdlebaugh and Miller 1980). At 26°C, ammonia nitrogen removal was lower than that observed at 13°C.

The seasonal performance data for the full scale RBC plant at Gladstone, Michigan (Hynek and Iemura 1980), indicated that only ammonia nitrogen removal was affected by the temperature. At 9.4°C (43°F), 67.3 percent ammonia nitrogen removal was reported, and removals increased to 93 percent at 17.2°C (63°F).

Given (1980), investigating the performance of RBC (carbonaceous substrate removal) units treating dilute wastewater, reported that temperature was not a significant variable over the range of temperatures observed (5-11°C).

Scale up effects on RBC performance

Wilson et al. (1980) demonstrated that scale up factors are significant in RBC process performance. The two RBC systems used by Wilson et al. were 0.5 m (1.64 ft) and 2 m (6.56 ft) in diameter, and the volume to surface ratios were 4.83 l/m^2 for the large scale and 5.53 l/m^2 for the small scale. Peripheral velocities of both units were 18 m/min (59 ft/min). In addition both units were loaded at the same hydraulic and organic loading rates using a screened municipal wastewater. Carbonaceous substrate removal (measured as weight of COD per discs area per day) in the small scale was 16 percent higher than the removal observed in the larger scale unit. There was no significant difference in TKN removal per unit area between the two sizes. The lower COD removal observed with the larger unit was attributed to less aeration capacity. The DO concentration in the first stage of the larger unit was 1.9 mg/l, while the smaller size contained 3.9 mg/l in the first stage.

Several authors (Bintanja et al. 1975; Ouano 1978; Kim and Molof 1980) developed oxygen mass transfer models relating the operational conditions and physical characteristics of the RBC system. Ouano developed the following relationship for the oxygen mass transfer coefficient constant:

$$K_L = D_L \frac{A_t}{V_t} \left(\frac{D^2 \zeta}{\mu} \omega\right)^{0.59} \quad (1)$$

in which

K_L = oxygen mass transfer coefficient, cm/sec
D_L = diffusivity of oxygen in water, cm^2/sec
A_t = surface area of tank, cm^2
V_t = volume of tank, cm^3
D = diameter of disc, cm
ζ = water density, g/cm^3
μ = water viscosity, g-cm/sec
ω = angular velocity, rad/sec

Kim and Molof (1980) showed that the volumetric oxygen transfer coefficient, $K_La(1/t)$, is proportional to ω and D as shown below:

$$K_{La} = K \cdot \omega^{1.07} D^{0.39} S^{-0.86} \quad (2)$$

where

S = half space between discs, L
K = constant
K_La= volumetric oxygen transfer coefficient, t^{-1}

Kim and Molof (1980) conducted their experiment with different values of ω, D and S, while Ouano changed only the rotational speed ω.

Equation 2 indicates that scaling factors should consider angular velocity as well as peripheral velocity for systems that might experience high organic loading rates. Friedman et al. (1979) indicated that the kinetic constants for COD removal were correlated with the log of angular velocity.

Olem and Unz (1980) discussed scale up from the point of view of the hydraulic characteristics obtained with two sizes of RBC units (0.5 m and 2 m diameters). A hydraulic analysis of the four stage units showed that for the smaller size the hydraulic characteristics were 78 percent mixed flow, 20 percent plug flow, and 2 percent dead space. The larger unit hydraulic performance was poorer than that exhibited in the smaller unit: 80 percent mixed flow, 10 percent plug flow, and 10 percent dead space.

Attached Biomass Characteristics

Microscopic observations and microbiological examinations of the RBC attached biomass have revealed a succession of life forms through the stages. Several studies (Torpey et al. 1971; Pretorius 1971; Pescod and Nair 1972; Antonie and Welch 1969; Hoag et al. 1980) have reported the presence of thick gray (or black with white patches) filamentous growth in the first stage, followed by thinner brownish-tan growth in the following stages, with patchy growth in the last stages. The brownish-tan growth is associated with nitrifiers. The predominant organisms present on all discs were reported to be Sphaerotilus and zoogleal bacteria. The growth on the first few stages is predominantly filamentous organisms (Sphaerotilus, Geotrichum) and the autotrophic sulfur bacteria Beggiatoa. Free swimming protozoa also were found in the first stages. Rotifers, stalked ciliates, nematodes, and fungi dominated the growth of microorganisms in the last stages. A recent study on RBC microfauna (Hoag et al. 1980) reported that the population levels of the microfaunal groups and shifts of the microfaunal groups were related to the COD and NH_4-N concentrations.

A conceptual dynamic model of the microbial film process was developed by Characklis and Trulear (1980). The major process steps were described as organic adsorption at the wetted surface, transport of the microorganism to the wetted surface, attachment of microorganisms by the production of extracellular polymers, metabolism and growth, and detachment of biofilm by fluid shear forces. Both sloughing and the amount of attached biomass were reported to be dependent on the influent organic loading rate. Pescod and Nair (1972) estimated that the wet weight of biomass was 0.7 $kg/m^2/d$, 1.1 $kg/m^2/d$ and 1.7 $kg/m^2/d$ at organic loading rates (expressed as kg $COD/m^2/d$) of 0.006, 0.012 and 0.024, respectively.

Mikula (1972) showed that the quantity of attached biomass could be predicted using a saturation-type curve from the organic loading rate. The rate of sloughing was unpredictable. Characklis and Trulear (1980) showed that sloughing increased with biofilm thickness. The sloughing mechanism has not been established; however, it has been suggested that shearing forces, starvation conditions, and anaerobic conditions within the attached biomass cause sloughing. Ouyang (1980) showed that the variation of attached biomass with BOD_5 loading was described by a saturation type relationship. Maximum values of dry weight of 65 $g/m^2/d$ were reported.

Kornegay and Andrews (1968a) worked with an annular reactor, rotating at 100 rpm. A glucose solution was used as the influent substrate, and it was observed that at a biofilm thickness greater than 70 μ the substrate removal rate did not increase. This thickness was defined as the active thickness of the biofilm, and the active thickness did not vary with the dissolved oxygen concentration over a range of 2.5 to 18 mg/l. Kornegay and Andrews (1968b) also reported that an increase in the rate of rotation decreased the overall film thickness from 300 μm at 50 rpm to about 100 μm at 200 rpm. The films were more uneven at the lower rotational speeds. The reported density was 95 mg/cm^3 on a dry weight basis. When a film thickness of 100 μm to 250 μm was maintained on a horizontal cylinder-attached growth system rotating at 2 rpm and treating a synthetic substrate, the nutrient utilization rate stabilized and did not increase as the thickness increased beyond 200 μm (Hoehn and Ray 1973). The film density increased to a maximum value of 105 mg/cm^3 at thicknesses of 100 μm to 150 μm, then decreased, and beyond 400 μm stabilized at about 25 to 30 mg/cm^3.

The effect of biofilm thickness on substrate removal rate was shown by Weng and Molof (1980) using a six-stage, 15 cm (6 inches) RBC unit, fed with synthetic substrate. The active film thickness was calculated to be between 135 to 265 μm, and beyond that thickness substrate removal did not increase. However, a decrease in COD removal occurred when the film thickness was between 350 μm and 600 μm. Substrate removal was relatively stable when film thickness was above 600 μm.

Chemical composition of the biomass as reported by Kornegay and Andrews (1968b) was 42.8 percent carbon, 7.6 percent hydrogen, and 9.95 percent nitrogen based on the volatile fraction. The ash content was 10 percent. Mikula (1979) reported a nitrogen content of 6 to 8 percent in the first two stages and 2 to 5 percent in the third and fourth stages, based

on total solids. Using primary municipal wastewater in RBC studies, Ouyang (1980) reported a carbon content of 35.2 to 39.1 percent, hydrogen 5.4 to 6.7 percent, nitrogen 5.8 to 6.9 percent, oxygen 21.4 to 25.1 percent, and ash 23.7 to 32.2 percent. An empirical chemical formula for the biomass (volatile solids base) was calculated by Ouyang (1980) to be $C_{4.2}H_{8.1}N_{0.6}O_{2.1}$.

Kinetic Studies and Predictive Models

Carbonaceous substrate removal

The existing mathematical models for steady-state carbonaceous substrate removal in an RBC fall into three general categories: a) statistical predictive models based on multiple regression of system and operational parameters, b) apparent reaction and growth kinetic models, and c) fundamental models that incorporate simultaneous mass transport of substrate and oxygen with growth kinetics.

Statistical models. Joost (1969) introduced a conceptual model for BOD reduction per stage as follows:

$$\frac{\text{BOD reduction}}{\text{Stage}} = K\, C^a\, R^b\, T^c\, S^d \qquad (3)$$

where

K = treatability factor
C = substrate concentration in the stage
R = physical configuration constant of RBC
T = temperature
S = reaction residence time

Wu et al. (1980), using available RBC data, derived the following predictive model for soluble BOD_5 removal:

$$F = \frac{14.2\, q^{0.5579}}{e^{0.32N}\, L_o^{0.6837}\, T^{0.2477}} \qquad (4)$$

where

F = fraction of influent soluble BOD_5 remaining in the effluent
q = hydraulic loading, gpd/sq ft
N = number of stages
L_o = influent soluble BOD_5, mg/l
T = temperature, °C

Apparent reaction and growth kinetics. Kornegay and Andrews (1968a) developed a kinetic model based on Monod growth kinetics for carbonaceous substrate removal for a fixed film system. Data collected with a rotating drum reactor using glucose as substrate were used to develop the following model:

$$F(S_o - S_1) = \frac{\hat{\mu}}{Y} A_w\, X\, d\, \frac{S_1}{K_s + S_1} \qquad (5)$$

in which

F = flow rate, L^3/t
S_o = influent substrate concentration, M/L^3
S_1 = effluent substrate concentration, M/L^3
$\hat{\mu}$ = maximum specific growth rate, 1/t
Y = cell yield coefficient
A_w = effective surface area, L^2
X = attached biomass concentration, M/L^3
d = active depth of the biofilm, L
K_s = half saturation constant, M/L^3

The temperature was controlled at 25°C. The following kinetic parameters were calculated: yield coefficient of 0.26 g volatile solids (VS)/g glucose, maximum specific growth rate, $\hat{\mu}$, of 0.28/hr (6.72/d), and a half-saturation constant, K_s, of 121 mg/l glucose. The thickness of the active layer was determined to be 70 μm.

Clark et al. (1978) applied the Kornegay and Andrews (1968a) model to predict soluble BOD_5 reduction with a four-stage, 2 m diameter, pilot plant RBC, fed with domestic wastewater primary effluent. The model was expanded by calculating individual kinetic constants for each stage except for the yield coefficient. The active attached biomass in each stage was estimated to be 70 percent of the total attached biomass. The kinetic coefficients obtained by Clark et al. are summarized in Table 1. The stage data showed that particulate BOD_5 removal and soluble BOD_5 removal occurred.

Mikula (1979) determined the temperature effect on maximum specific growth rate using an Arrhenius type relationship. The model included the biomass in suspension, and the model for the first stage was:

$$F(S_o - S_1) = \frac{A_r e^{-E_a/RT}(A_w X_f + V X_s)}{Y} \left[\frac{S_1}{K_s + S_1}\right] \qquad (6)$$

in which

A_r = frequency factor, t^{-1}
E_a = activation energy, cal/mol
R = universal gas constant = 1.987 cal/mol°K
T = absolute temperature, °K
X_f = total attached biomass, M/L^2
X_s = suspended biomass, M/L^3
V = stage volume, L^3

The other terms have been defined previously. Mikula used a four-stage, 1.2 m (4 ft) diameter, RBC pilot plant, to treat a cheese processing wastewater. Frequency factors of 1.4×10^{12} to 1.5×10^{12}/d were obtained for the first two stages, and lower values, 0.26×10^{12} to 0.46×10^{12}/d for the last two stages. The activation energy was found to be almost the same for all four stages, 17,780 to 17,850 cal/mole. The half-saturation constants were 32 mg/l for the first stage, 186 mg/l for the second, and 59 mg/l for the third stage. Difficulties were reported in determining the fourth stage value. All of the constants were based on soluble COD.

An RBC model has been developed by several investigators using first order kinetics; however, Monod kinetics have not been used. A first order kinetic model for an RBC treating industrial wastewaters was developed by Eckenfelder and Vandevenne (1980)

$$\frac{S_n}{S_o} = \frac{1}{(1 + KA/Q)^n} \qquad (7)$$

Table 1. Kinetic coefficients determined by Clark et al. (1978) (based on soluble BOD_5).

Stage	Apparent Yield Coefficient y_a	Active Biomass X_a, g/m^2	Half Saturation Constant K_s, mg/l	Maximum Specific Growth Rate $\hat{\mu}$, day^{-1}
1	0.96	35.8	431	4.4
2	0.96	30.6	546	3.8
3	0.96	24.4	32	0.2
4	0.96	9.8	8	0.3

in which

S_n = effluent soluble BOD_5, mg/l
S_o = influent soluble BOD_5, mg/l
Q = flow rate, L^3/t
A = effective surface area, L^2
K = proportionality constant
n = number of stages

The model was verified with available data, and in some cases in the later stages, the K value decreased. The model also was used to predict the performance of an RBC treating domestic wastewaters. The K values were found to be in the range of 0.204 to 0.701 $m^3/m^2/d$. Kincannon and Groves (1980) explained the performance of a laboratory scale, four-stage, 15 cm (6 inch) diameter RBC treating glucose (COD of 300 mg/l) by using first order kinetics and incorporating the suspended biomass into the expression

$$\frac{(S_i - S_e)F}{X_m + VX_s} = 0.13\, S_e \qquad (8)$$

in which

S_i = influent substrate, mg/l
S_e = effluent substrate, mg/l
X_m = active biomass (taken as a constant for each stage), mg
V = liquid volume of stage, liter
X_s = suspended growth, mg/l
F = flow rate, l/d

Benjes (1978) applied two types of equations similar to trickling filter design equations to analyze RBC plant data

$$\frac{L_e}{L_o} = e^{-K(V/Q)^{1/2}} \qquad (9)$$

$$\frac{L_e}{L_o} = e^{-S(A_s/Q)^{1/2}} \qquad (10)$$

in which

L_e = effluent BOD_5, mg/l
L_o = influent BOD_5, mg/l
V = reactor volume, L^3
A_s = surface area, L^2
Q = flow rate, L^3/t
K,S= constants

Using data from Pewaukee, Wisconsin, and Gladstone, Michigan, to verify Equations 9 and 10, Benjes (1978) recommended a design value for S at 20°C of 0.055 for BOD_5 removal with an RBC. Given (1980) reported a second order reaction equation which described BOD removal better than zero or first order equations. However, the best fit of the performance data was obtained using a multiple regression analysis relating the effluent BOD and the influent organic loading rate.

Mass transport models. Dynamic and steady state models for carbonaceous substrate removal with an RBC were developed by Grieves (1972). Two types of models were developed: a) a pseudo homogeneous model, and b) a heterogeneous model. The pseudo homogeneous model consists of external mass transfer resistance (in the attached liquid film), and the heterogeneous model consists of external and internal (diffusion into biofilm) mass transfer resistance. The substrate consumption rate was considered to obey Monod growth kinetics. For first order Monod kinetics, the model for an n stage reactor is:

$$\frac{C_b}{C_o} = \frac{1}{1 + \frac{N}{F}\{P_1 A_s + P_2 W[1 - \exp(\frac{-P_1 A\alpha}{P_2 W})]\}} \qquad (11)$$

in which

C_b, C_o = substrate effluent and influent, mg/l
N = total number of discs per stage
F = influent flow rate, L^3/T
A_s = submerged area of the disc, L^2
W = rotating speed, rpm

$$P_1 = \frac{K_L K_1}{1 + K_1}$$

$$K_1 = \frac{\hat{\mu} \cdot X_a \cdot d(TF)^{n-1}}{YK_s \eta K_L}$$

$$P_2 = K_2 \pi [R_1^2 - R_2^2] \Delta L$$

$A\alpha$ = unsubmerged area of the disc, L^2
K_L = mass transfer coefficient, L/t
TF = treatability factor
n = number of stages
K_2 = 1, assuming no dripping of liquid film
R_1, R_2 = radius of disc and of unsubmerged circle, L
Δ = liquid film thickness, L
η = effectiveness factor

The other terms were defined previously. The dynamic model was calibrated by using dynamic test data obtained using glucose as the substrate, and the results predicted with the calibrated model agreed with data reported by Torpey et al. (1971).

Paolini et al. (1979) used a three-stage, 27.5 cm diameter, bench scale RBC unit to treat an industrial wastewater. The mathematical model developed by Grieves (1972) was used to evaluate the experimental results. Organic loading rates in the range of 8.7 to 27.7 g$BOD_5/m^2/d$ resulted in diffusion governed kinetics for rotational speeds less than 0.6 rpm. At higher rotational speeds, a Monod equation based model such as the Kornegay model (Kornegay and Andrews 1968b) provided a good approximation.

Schroeder (1977) presented a diffusion-reaction model for an RBC process. The model was based on diffusion control without external mass transfer resistance. Also the model assumed constant active biomass. The proposed model was:

$$M_z = \frac{f\ h\ k_o^*\ C_b^2\ A_s}{K_m + C_b} \qquad (12)$$

in which

M_z = mass of substrate removed per unit time, M/t
f = proportionality factor, M/L^3
h = biofilm thickness, L
k_o^* = $k_o X$
X = biomass concentration, M/L^3
k_o = reaction constant, 1/t
A_s = submerged area of the discs, L^2
C_b = bulk substrate concentration, M/L^3
K_m = constant

Friedman et al. (1976) modified the Schroeder model and simplified it to the following:

$$M_z = \frac{K' A_s C_b^2}{K_m + C_b} \qquad (13)$$

where

M_z = mass of substrate removed per time, M/t
A_s = surface area, L^2
C_b = bulk substrate concentration, mg/l
K', K_m = constants

To simulate a continuous removal function, the plug flow approximation was applied by Friedman et al. (1976) to obtain the following equations:

$$K_m \left[\frac{1}{C_b} - \frac{1}{C_{bi}}\right] + \ell n \left[\frac{C_{bi}}{C_b}\right] = \frac{K' A_s}{V}\theta \qquad (14)$$

$$\frac{K' A_s}{V}\theta = \frac{K''\theta}{V} \qquad (15)$$

in which

C_{bi} = influent substrate concentration, mg/l
θ = hydraulic detention time per surface area, t/L^2
V = reactor volume per surface area, L^3/L^2
K'' = constant

It was reported that experimental data collected at 20°C using a synthetic substrate (sucrose) resulted in mean K_m values of 22.3 mg/l COD. K" values were related to the initial substrate concentration and detention time as shown in the following relationship

$$K'' = k[\theta C_{bi}]^{0.664} \qquad (16)$$

in which

k = constant

Friedman et al. (1979) analyzed additional experimental data and reported K_m values of 42.9 mg/l COD ± 31.4 mg/l COD. A relationship between the K" values and the rotational speed was also presented:

$$K'' = [a\ \ln(\theta C_{bi}) + b]\ \ln W \qquad (17)$$

in which

W = rotational speed, rpm
a, b = constants

A second order nonlinear ordinary differential equation model was developed by Williamson and McCarty (1976) for substrate utilization within biofilms with external and internal mass transfer resistance using Monod growth kinetics. The equation was solved using a Runge Kutta finite difference technique for several simulated conditions.

Rittmann and McCarty (1978, 1980), using the concepts developed by Williamson and McCarty (1976), showed that substrate removal kinetics could be expressed as variable order kinetics with reaction orders of $0.5 \leq q \leq 1.0$ where q is reaction order. The model was verified using constants from the literature and the data obtained by Bintanja et al. (1976).

The mass transport models were expanded by Harris and Hansford (1976) by incorporating oxygen mass transport with substrate transport and substrate oxygen limiting reactions as follows:

$$\frac{d^2S}{dz^2} = \frac{\hat{\mu} x}{YD_s}\left(\frac{S}{K_s+S}\right)\left(\frac{O}{K_o+O}\right) \qquad (18)$$

$$\frac{d^2O}{dz^2} = \frac{\hat{\mu} x F}{YD_o}\left(\frac{S}{K_s+S}\right)\left(\frac{O}{K_o+O}\right) \qquad (19)$$

in which

S = substrate concentration, M/L^3
O = oxygen concentration, M/L^3
z = distance measured into the slime, L
x = attached biomass concentration, M/L^3
K_s, K_o = half-saturation constants for substrate and oxygen, M/L^3
D_s, D_o = diffusion constants for substrate and dissolved oxygen, L^2/t
F = constant factor relating quantities of substrate and oxygen utilized in the aerobic metabolism process

The solution of Equations 18 and 19 for the kinetic constants required trial and error procedures using numerical techniques. Substrate removal rates were measured using a synthetic waste, and the biomass was scraped from the RBC daily. A mixed liquor dissolved oxygen concentration of 6-8 mg/l was maintained. Oxygen limiting conditions were indicated with influent COD concentrations of 500 mg/l or bulk concentrations in the reactor of 110 mg/l. The calculated active biomass depth was about 70 µm. Bulk COD concentrations below 75 mg/l resulted in substrate limiting conditions, and at these lower bulk concentrations the calculated active biomass depth ranged from 70 to 90 µm.

Famularo et al. (1978) applied mass transport concepts, including oxygen mass transfer, to RBC systems. The model was applied in the same manner as the previous mass transport models The model fit RBC data from paper mill industries and domestic sources regardless of which reaction order was simulated (zero order or first order).

Ammonia nitrogen removal

Weng and Molof (1974) used a six-stage, 15 cm (6 inch) diameter, RBC unit fed with synthetic substrate to investigate the nitrification process in RBC units. A multiple regression model was developed to describe the results of the experiments

$$F = \frac{0.0545\ L_o^{0.644}\ Q^{0.414}}{S^{0.53}\ A^{1.276}} \qquad (20)$$

in which

F = fraction of influent NH_4-N remaining in the effluent
L_o = influent NH_4-N, mg/l
Q = flow rate, l/day
S = rotating speed, rpm
A = discs area, sq ft (0.09 sq m)

Using nitrification data collected at the Pewaukee, Wisconsin, RBC plant and pilot RBC data, Wu et al. (1980) used multiple regression analyses to develop two models. One model was developed for influent soluble BOD_5 concentrations greater than 80 mg/l to simulate secondary treatment and the other model was for influent soluble BOD_5 concentrations less than 50 mg/l, simulating tertiary treatment. The secondary treatment model for a four-stage RBC was as follows:

$$F = \frac{4.336\ q^{0.1692}}{L_o^{0.2395}\ T^{0.6229}} \qquad (21)$$

The tertiary treatment model for a four-stage RBC resulted in the following equation:

$$F = \frac{1.552\times10^{-4}\ q^{0.6803}\ L_o^{2.6452}}{T^{0.217}} \qquad (22)$$

in which

F = fraction of influent NH_4-N remaining in the effluent
q = flow rate, gpd
L_o = influent NH_4-N, mg/l
T = temperature, °C

Watanabe et al. (1980) presented a mass transport model with a simultaneous zero order reaction within the active biofilm. Consequently, the overall reaction rate appears to be first order in low level bulk NH_4-N concentrations, one-half order within medium concentrations, and zero order with high bulk concentrations of NH_4-N. Using data collected from 30 cm diameter, RBC units, fed with synthetic substrate, the following relationships were developed:

$$K^* = (0.18 - 0.21)\ C_b^{-(0.82 - 0.90)} \qquad (23)$$

$$K^* = \left(\frac{1}{K_d} + \frac{1}{K_r}\right)^{-1} \qquad (24)$$

$$-V_r = K^*\ C_b \qquad (25)$$

where

K^* = overall rate constant, m/hr
C_b = ammonia-N concentration in the bulk liquid, mg/l
K_d = mass transfer coefficient (liquid layer), m/hr
K_r = biofilm coefficient (diffusion-reaction lumped parameter), m/hr
V_r = nitrification rate, g/m^2/hr

A mass transport model coupled with Michaelis-Menten enzyme kinetics to describe substrate and oxygen limiting conditions was developed by Mueller et al. (1980). The model was calibrated using RBC data from Gladstone, Michigan, and the following kinetic parameters resulted:

Maximum specific growth = 0.045/d;
Half-saturation constant for NH_4-N = 2.0 mg/l;
Half-saturation constant for dissolved oxygen = 0.5 mg/l;
Decay rate = 0; and
Yield coefficient = 0.17.

Monod growth kinetic relationships were used by Saunders et al (1980) to describe the performance of 25.4 cm diameter, RBC units treating synthetic substrate and controlled for mean cell residence time of the attached growth. The following results were reported: maximum specific growth rate = 0.56 to 0.83/d and the half-saturation constant = 0.18 to 1.0 mg/l Ito and Matsuo (1980) worked with a 25 cm diameter RBC processing a synthetic substrate and reported maximum ammonia nitrogen removal of 4 g/m^2/d, a yield coefficient of 0.15, and a maximum specific growth rate of 0.03/d.

Temperature effects on the kinetic constants

Carbonaceous substrate removal. The following expression showing the temperature dependence of BOD_5 removal efficiency was reported by Ellis and Banaga (1976):

$$E_{T1} = E_{T2}\ \theta^{T1-T2} \qquad (26)$$

in which

E_{T1} = efficiency at temperature T1, °C
E_{T2} = efficiency at temperature T2, °C
T = temperature, °C
θ = temperature factor

At temperatures between 11°C and 18°C, θ was found to be 1.006 and for temperatures between 18°C and 27°C, θ was 1.002. Conducting experiments with cheese processing wastewater, Mikula (1979) found that the variation of the maximum specific growth with tempera-

ture could be expressed by an Arrhenius type relationship

$$\hat{\mu} = A_r e^{-E_a/RT} \quad (27)$$

in which

A_r = frequency factor, 1/t
E_a = activation energy, cal/mole
R = universal gas constant
T = absolute temperature, °K

The values of A_r were calculated to be 1.4×10^{12} to 1.5×10^{12}/day for the first two stages and 0.26×10^{12} to 0.46×10^{12}/day for the last two stages. The activation energy varied from 17,780 to 17,850 cal/mole.

Ammonia nitrogen removal. The effect of temperature on the nitrification rate in an RBC assuming zero order nitrification was reported by Watanabe et al. (1980) to be:

$$(rd)_T = (rd)_{20} (1.05)^{T-20} \quad (28)$$

in which

$(rd)_T$ = nitrification rate constant 1/hr at temperature T
$(rd)_{20}$ = nitrification rate constant 1/hr at 20°C
T = temperature, °C

The expression was developed with data collected in experiments conducted at temperatures of 15°C, 23.5°C, and 28.5°C. Mueller et al. (1980) calibrated a model using the following expression to relate temperature and maximum specific growth rate (1/day)

$$\mu_T = \hat{\mu}_{20}(1.1)^{T-20} \quad (29)$$

Murphy et al. (1977) reported that nitrification dependence on temperature follows the relationship:

$$K = 9.45 \times 10^{11} e^{-13,900/RT}; \text{ for } 7°C \text{ to } 25°C \quad (30)$$

in which

K = reaction rate constant, 1/hr
R = universal gas constant, cal/g mole/°K
T = temperature, °K

Summary

When used to treat municipal wastewater, RBC systems achieve 85 to 95 percent carbonaceous substrate removal along with substantial nitrification. While the major part of the carbonaceous substrate is removed from the wastewater in the first stage, the succeeding stages polish the carrier stream, stabilize the sloughed biomass, and provide for nitrification. The succession of life forms and the changes in appearance of the attached biomass as the wastewater flows through the stages correspond with the functional characteristics of the stages and the characteristics of the carrier stream. At tank volume to surface area ratios greater than 4.9 l/m^2 (0.12 gal/ft), performance is not improved.

The rotational speed, among other factors, affects the aeration capacity of the system. High rotational velocities increase energy consumption, stimulate the sloughing of attached biomass, and might increase sludge production. Generally, a rotational speed corresponding to a peripheral velocity of 18 m/min is considered optimum. Some oxygen transfer studies and scaling studies indicated that the optimum design should consider also the angular velocity. Scaling factors appear to be important with regard to hydraulic characteristics and performance.

The carbonaceous substrate removal efficiency with an RBC treating domestic wastewater generally increases with higher influent substrate concentrations and higher temperatures, and decreases with higher hydraulic loading rates. In some cases, temperature increases did not improve carbonaceous removal or nitrification. The low DO concentrations in the mixed liquor at higher temperatures were found to be the cause for poorer performance. Controlled temperature studies on RBC performance are limited, and generally represent full scale operational data. Most designs consider only temperatures below 13°C (55°F); however, there are significant differences in temperature safety factors (Chesner and Ionnone 1980) employed in several design methods.

Extensive performance studies have been conducted on pilot and full scale plants, and in some studies the performance data represent only the characteristics of the final effluent from the RBC and the secondary clarifier. Very little stage by stage data are available. The existing data from studies treating municipal wastewater generally indicate that the kinetics for carbonaceous substrate removal are first order with substrate limiting phenomena. Nitrification kinetics were reported to be zero or first order.

Several studies (Kornegay and Andrews 1968b; Clark et al. 1978; Mikula 1979) employed Monod kinetics to describe carbonaceous substrate removal in fixed film reactors. Kornegay and Andrews (1968b) used a rotating drum, and the others an RBC. Using domestic wastewater and soluble BOD_5 as the measure of performance, Clark et al. (1978) reported values of 0.96 for the yield coefficient, 4.4/d for the maximum specific growth rate in the first stage, and 431 mg/l for the half-saturation constant in the first stage. The Monod equation was used by Saunders et al. (1980) to model nitrification in the RBC. The half-saturation constant ranged between 0.18 and 1.0 mg/l, and the range of values for the maximum specific growth rate was 0.56 to 0.83/d.

Data showing the temperature dependence of kinetic constants in RBC systems are limited. Murphy et al. (1977) developed a relationship between temperature and the reaction rate constant using an Arrhenius type relationship. Mikula (1979) used the Arrhenius relationship to relate temperature and the maximum specific growth when treating a cheese processing wastewater. Ellis and Bananga (1976) developed a relationship between BOD_5 removal efficiency and temperature for temperatures ranging between 11°C and 27°C.

Mass transport models incorporating active biofilm depth and external and internal mass transfer resistance with simultaneous substrate removal reactions generally used Monod type kinetics. The models consist of second order differential equations amenable to solution by numerical methods using

estimated values of mass transport and kinetic constants. Schroeder (1977) presented a mass transport model for pseudo-homogeneous biofilm configurations incorporating simultaneous diffusion and reaction. Kinetic studies using a synthetic substrate in an RBC unit employed Schroeder's (1977) model to analyze the data (Friedman et al. 1976; Friedman et al. 1979).

Some of the models describing carbonaceous substrate removal were based only on soluble influent substrate, although the RBC data in most cases indicate that there is also particulate substrate removal in an RBC.

Presently the approach to the design of the RBC process is empirical, because a reliable substrate removal kinetic model for a wide range of temperatures has not been developed. Although several kinetic models have been proposed, they have limited applicability because of the limitations of the data used to develop the models. Uncontrolled temperature conditions, variations in hydraulic and organic loading rates, and DO deficiency conditions are examples of the difficulties encountered by many researchers. Some of the existing models are based on experiments conducted with synthetic substrate and others based on theoretical considerations with little experimental verification.

The unknown features of the RBC process include:

1) Substrate removal kinetics at each stage.
2) Reproducible values for the kinetic constants.
3) Effects of temperature on the kinetic constants.
4) Dynamics of biomass growth, decay, and sloughing.
5) An oxygen balance for the RBC system considering system aeration capacity and oxygen consumption for exogenous substrate removal, endogenous respiration, and nitrification.

To develop rational design criteria, the RBC process must be investigated under controlled experimental conditions for each category of wastewater. The research has to focus on:

1) Developing a substrate removal kinetic model for each stage.
2) Determining the kinetic constants of substrate removal.
3) Determining the dynamics and the constants of biomass growth decay and sloughing for each stage.
4) Determining the effect of temperature on the kinetic constants.
5) Performing an oxygen balance.
6) Developing design criteria for the RBC process under aerobic conditions for a wide range of temperatures.

MATERIALS AND METHODS

Experimental Design

The experimental RBC units were operated continuously at constant influent flow rates, constant wastewater, percentage and constant temperature. The influent wastewater was maintained at a constant temperature in a refrigerated storage tank, and the experimental RBC units were located in a constant temperature room to maintain the desired water temperature through the four stages of the RBC units.

Although wastewater has uncontrolled fluctuations in substrate concentration, its use was preferred over a synthetic substrate because of the difficulty in correlating synthetic substrate experimental data with field data collected at RBC units treating domestic wastewater. To minimize fluctuations in the concentration of pollutants in the wastewater applied to the experimental units, batches of wastewater were collected from the wastewater treatment plant in Hyrum, Utah, and stored in a refrigerated tank. The wastewater was collected at the same time of day during the weekdays to ensure maximum uniformity in wastewater composition and strength through each phase of the study. The wastewater was diluted with tap water to provide four strengths of wastewater for application to the RBC units.

Kinetic constants for carbonaceous substrate removal in the RBC process were determined using steady-state chemical oxygen demand data collected from the four RBC bench scale units operating simultaneously at the same hydraulic loading rate and temperature.

Four organic loading rates were used in each of the three constant temperature phases of the project. The organic loading rates were maintained by diluting the raw wastewater with dechlorinated tap water.

To determine the dependence of the kinetic constants on temperature, the research was conducted in three phases at three different temperatures. Each temperature phase was started with the RBC units free of biomass. This was done to avoid the influence of earlier cultures grown at different temperatures on the experiments.

Wastewater Source

The wastewater used in this study was collected at the Hyrum City wastewater treatment plant, Hyrum, Utah. Hyrum City has a population of 2500 and is located in a located in a rural area 16 km (10 miles) south of Utah State University, Logan, Utah. The wastewater flow rate into the treatment plant ranges from 1500 to 3400 m^3/d (0.4-0.9 mgd) and the BOD_5 concentration ranges from 70 to 140 mg/l. The treatment plant consists of screening, an oxidation ditch, secondary clarifier, sand filters, and chlorination. Sludge is processed with aerobic digestion and sludge drying beds.

Comminuted wastewater was collected for use with the RBC units and hauled to the laboratory. A 1.5 HP centrifugal pump[1] was used to pump 2.3 to 3.0 m^3 (600-800 gal) of wastewater into a tank truck. Wastewater was collected three times a week (on Mondays, Wednesdays, and Fridays) between 10:30 a.m. and 11:00 a.m. The selection of this schedule was based on a survey conducted at the Hyrum plant before the laboratory study was started. The survey showed that peak values of COD concentration (~ 300 mg/l) occurred between 10:30 a.m. and 11:00 a.m. The wastewater was transported to the Utah State University campus and pumped to a refrigerated storage tank located on the second floor of the Engineering Building. A 5 cm (2 inch) diameter PVC pipeline was installed in the Engineering Building for this purpose. To avoid pumping solids that settled during transportation, the wastewater was pumped from the tank truck through a ball valve installed 30 cm (~ 1 ft) above the tank floor. The residue was drained from the tank truck through another ball valve installed at the level of the tank floor. The settled wastewater was stored in the laboratory in a 1.9 m^3 (500 gal) stainless steel tank. The tank was equipped with a mixer and a thermostatically controlled cooler that was used to maintain the temperature of the wastewater at 2°C. During the weekends an additional plastic storage tank with a capacity of 0.76 m^3 (200 gal) was used to provide the volume of wastewater required for 3 days of operation. Approximately 0.76 m^3/d (200 gpd) of wastewater was required to operate the RBC units continuously.

Influent Feeding System

Wastewater was pumped continuously into a 26.7 x 29 x 29 cm plexiglass constant head tank located in the constant temperature room. The wastewater flow rate into the wastewater heating chamber was regulated with the constant head tank. The overflow from the constant head tank drained by gravity back to the cold storage tank. Culinary water at a temperature of approximately 15°C was filtered through a 7.5 cm (3 inch) diameter by 30 cm (1 ft) activated carbon column and discharged into the water heating chamber. During the third phase of the study, conducted at 5°C, it was necessary to cool the tap water to the required temperature. The tap water was cooled by passing the water through a coil of 1.25 cm (0.5 inch) diameter copper tubing 18 meters (60 ft) in length submerged into the refrigerated wastewater storage tank. The tap water that passed through the coil was cooled to 2-3°C, and then flowed through the activated carbon column to the water heating chamber.

Both of the plexiglass heating chambers were 20 cm deep, 18 cm wide, and 20 cm long and contained

[1]Berkeley Pump Company, Berkeley, California.

about 9 liters of wastewater at the overflow level. The heating chambers drained by gravity and were equipped with mixers rotated by gear motors[2] at 12 rpm. Mercury thermoregulators[3] were used to control the 500 watt heater coils[4] to provide the desired influent temperatures. During the second phase of the study conducted at 20°C, two heaters of 500 watts each were installed in the wastewater heating chamber. A ten channel peristaltic pump[5] and a two channel variable speed peristaltic pump[6] were used to pump the wastewater and tap water to the mixing chambers for dilution. Four plexiglass 10 cm (4 inch) diameter mixing chambers with magnetic mixing bars received the required flows of wastewater and tap water. The water levels in the mixing chambers were maintained at a height of 2 cm and the influent to the four RBC units flowed by gravity from the mixing chambers. A schematic drawing of the feeding system is shown in Figure 2.

Tygon tubing was used in the influent application system with the exception being the silicon[7] tubing used with the peristaltic pumps. The overflow pipes were 1.9 cm (3/4 inch) diameter PVC.

[2]Dayton Electric Manufacturing Company, Chicago, Illinois.

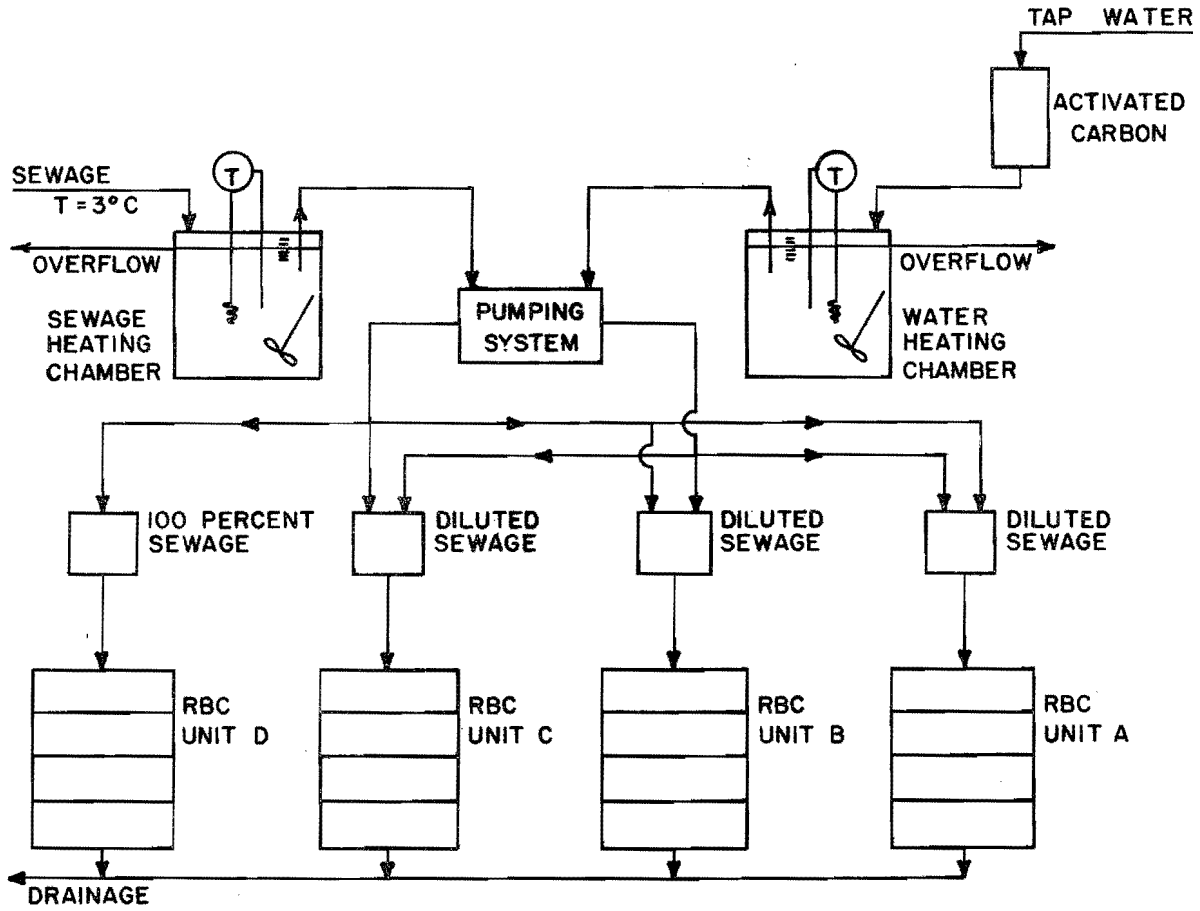
[3]Cole Parmer Instrument Company, Chicago, Illinois, Model 0-50°C.

[4]Technilab Instruments, Inc., Pequannock, New Jersey.

[5]Masterflex - Ten Channel Drives, Cole Parmer Instrument Company, Chicago, Illinois.

[6]Masterflex Pump Model 7545-00, with add on head; Cole Parmer Instrument Company, Chicago, Illinois.

[7]Catalog Number 6411, Cole Parmer Instrument Co., Chicago, Illinois.



Schematic of experimental apparatus employed to develop data base for developing kinetic constants.

Figure 2. Schematic diagram of experimental apparatus.

Experimental Rotating Biological Contactors

Four, four-stage, 38 cm diameter, bench scale RBC units[8] were employed. The disc media were composed of high density polyethylene strengthened with black carbon in a form expanded by a factor of 1.37. Each stage contained four 37.5 cm (14.8 in) diameter discs spaced 2.5 cm (1 inch) apart and two 22.9 cm (9 inch) diameter flat support side discs mounted on the shaft.

The total area was 1.375 m^2 (14.78 ft^2) per stage including the area of the side discs. The volume of the reactor vessel for each stage was 6 liters and the discs operated at a submergence of 33.3 percent.

During the first phase of the study, conducted at 15°C, all of the units were constructed of steel and powered by a fixed speed gearhead motor[9] operating at 16 rpm. Motor failures experienced during this phase necessitated changing the RBC units[10] (in the second and third phases) and changing the power to a 1/15 HP[11] variable speed motor. The new RBC units were constructed of PVC and had slightly different dimensions than the first ones. Units employed in the second and third phases were equipped with 39 cm diameter discs with a total area of 1.474 m^2 (15.85 ft^2) per stage.

Table 2 summarizes the dimensions of the RBC units employed during the three phases of the study. Flow was perpendicular to the shaft and passed from one stage to another through 1.9 cm (0.75 inch) pipe approximately 30 cm (1 ft) in length. Effluents from the fourth stage of all units flowed by gravity through 1.9 cm (0.75 inch) rubber hoses to a drain.

The four experimental units were identified by the letters A, B, C, and D. The influent organic loading rates to these units increased successively from unit A through unit D. In all experiments, unit D received undiluted wastewater. In the third phase of the study conducted at 5°C, unit A was designed to receive 25 percent wastewater. The low COD concentrations (173 mg/l) in the wastewater during this period resulted in filtered COD concentrations of 15 mg/l in the effluent from the first stage of unit A. Consequently, it was decided to change the dilution rate for unit A to 65 percent. Steady-state data collected in the other units started approximately two weeks after the change in dilution percentage for unit A. It was necessary to operate the units for more than one month after the dilution change to collect adequate data under steady-state conditions.

Table 2. Summary of the dimensions of the RBC experimental units.

Phase	1	2, 3
Parameter		
Number of stages	4	4
Number of discs/stage	4	4
Discs diameter, cm	37.5	39.0
Inflation factor	1.37	1.37
Side discs diameter, cm	22.9	22.9
Total surface area/stage, m^2	1.375	1.474
Water volume/stage, ℓ	6	7
Submergence, %	33.3	35.5
Rotational speed, rpm	16	16
Flow rate, ℓ/hr	12	12
Detention time per stage, min	30	35

Sampling Procedures

Nine ISCO[12] composite samplers were used in this study to monitor the influent wastewater and the effluent from each stage. On each sampling day, a sample of the influent wastewater was collected. Samples from the stages of the RBC units were collected on alternate days because of the limited number of automatic samplers. The eight stages of two units were sampled on one day, and the remaining eight stages were sampled the next day. The influent samples were taken from the wastewater heating chamber at approximately 5 cm below the overflow level. The effluent samples from each stage were collected at approximately 1.5 cm below the mixed liquor surface at the outlet of the stage. The same sampling points were used when grab samples were collected. Grab samples of the tap water were taken from the water heating chamber. The influent flows to the units were monitored at the inlet to the mixing chambers.

The composite samples were collected at intervals of 20 minutes for 24 hours. The samples were stored in 3.79 liter (1 gal) glass bottles placed in two refrigerators.

The sampling equipment and the refrigerators were located in the constant temperature room with the RBC units. During the second phase of the study conducted at 20°C, it was necessary to store the glass bottles in ice buckets to maintain the samples at a temperature below 4°C.

Twenty-four hour composite samples of the influent wastewater and the effluent from all of the stages were analyzed one to three times each week for chemical oxygen demand (COD), suspended solids (SS), volatile suspended solids (VSS), Kjeldahl nitrogen, ammonia nitrogen, nitrite nitrogen, and nitrate nitrogen. Grab samples of the influent wastewater, tap water, and effluents from all stages of the RBC units were collected a minimum of two times during the steady-state period and monitored for the pH value and

[8]Environmental System Division, George A. Hormel and Company, Coon Rapids, Minnesota.

[9]Dayton Electric Mfg. Co., Chicago, Illinois, Model 147147.

[10]Environmental System Division, George A. Hormel and Company, Coon Rapids, Minnesota.

[11]Baldor Electric Company, Fort Smith, Arkansas.

[12]ISCO Environmental Division, Lincoln, Nebraska.

dissolved oxygen (DO) concentrations. Because of the controlled conditions and the wastewater source (domestic), it was sufficient to collect only a few samples for DO and pH, to represent the environmental conditions in each stage. At the end of each phase, the total quantity of attached biomass on each stage of the RBC units was determined, and the percentage of volatile solids (VS) in the attached biomass was determined. In addition, the attached biomass was analyzed for total organic carbon (TOC), total nitrogen, and total phosphorus. During the steady-state period of the second phase of the study conducted at 20°C, a composite sample of the influent wastewater and the effluents from the RBC stages were analyzed for TOC and biological oxygen demand (BOD_5).

The influent wastewater flow rates and the tap water flow rates were monitored daily. In situ water temperatures at each stage were measured four to seven times a week. Frequent influent tap water grab samples were taken to ensure that a chlorine residual was not present.

Analytical Methods

The analytical methods employed in this study were conducted according to Standard Methods (APHA 1975) with the exception of the following methods: the ampule technique (Oceanography International 1978) was used to conduct both total and filtered COD; and the influent wastewater COD and the effluent total COD were measured using the high level COD standard ampule method. In the standard ampule method the COD is determined using a spectrophotometer at 600 nm to measure the concentration of the Cr (+3) ion. The effluent filtered COD was measured using the low COD value ampule method. The COD is determined using a spectrophotometer at 440 nm by measuring the concentration of the Cr (+6) ion. All COD tests were performed on three replicates.

BOD_5 tests were conducted by adding 0.5 mg/l allylthiourea to suppress nitrification. Otherwise, the test was performed according to Standard Methods (APHA 1975). TOC measurements were made using the potassium persulfate and phosphoric acid digestion method with a carbon analyzer[13] (Oceanography International 1970). Ammonia nitrogen concentrations were performed according to the indophenol method described in Standard Methods (APHA 1975). Kjeldahl nitrogen concentrations were determined with an automated procedure using a Technicon[14] Auto Analyzer II as described in Industrial Method No. 376-75W/B (Technicon Industrial Systems 1977). Nitrite nitrogen was measured using a procedure available with a Technicon Auto Analyzer II. The procedure is described in Industrial Method No. 161-71W (Technicon Industrial Systems 1973a). Nitrate nitrogen was measured with the Technicon Auto Analyzer II according to the Industrial Method No. 100-70W (Technicon Industrial Systems 1973b).

Suspended solids and volatile suspended solids were analyzed according to Standard Methods (APHA 1975) using 0.45 μ Whatman GF/C glass fiber filters. Attached biomass solids were determined by applying an aluminum foil cover over the disc containing the attached biomass and drying overnight at 103°C. The dried discs were cooled and weighed. Representative samples of dry biomass were removed from the disc for further analysis, before washing the discs to remove all of the biomass. The cleaned discs were then dried at 103°C, cooled, and weighed. The percentages of volatile solids in the attached biomass were determined according to procedures outlined in Standard Methods (APHA 1975).

Biomass TOC was measured using the same procedure used to determine the TOC in liquids. Biomass total nitrogen percentages were determined by an automated procedure (based on micro Dumas Method) using a Coleman[15] Model 29 Nitrogen Analyzer (Coleman Instruments 1968). Biomass total phosphorus was measured according to the persulfate digestion method described in Standard Methods (APHA 1975). The iodometric method was used to measure residual chlorine, and the azide modification of the Winkler method was used to determine the dissolved oxygen concentrations (APHA 1975). Electrical conductivity was measured with a Yellow Springs[16] electrical conductivity meter. The pH value was determined using a Beckman pH meter.[17] Mercury thermometers were used to measure the temperature. Flow rates were measured by collecting the flow in a graduated cylinder over a 1 minute period of time.

Filtered COD, BOD_5, and TOC were conducted on samples filtered through pre-washed and dried 0.45 μ Whatman GF/C glass fiber filters. The tests for filtered nitrogen species were performed on samples filtered through 0.45 μ Millipore filters. All of the analyses, except the in situ measurements, were performed at the Utah Water Research Laboratory, Logan, Utah.

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[13]Oceanography International Co., College Station, Texas.

[14]Technicon Industrial Systems, Division of Technicon Instruments Corp., Tarrytown, New York.

[15]Coleman Instruments, A Division of the Perkin-Elmer Corp., Maywood, Illinois.

[16]Yellow Springs Instrument Co., Inc., Yellow Springs, Ohio.

[17]Beckman Instruments, Inc., Fullerton, California.

RESULTS AND DISCUSSION

General

The four experimental rotating biological contactor (RBC) units were operated from late October 1979 until mid July 1980. The study was conducted in three consecutive phases at three different temperatures of 5°C, 15°C, and 20°C. Each phase was started with "clean" RBC units (without biomass). The influent flow rates to the RBC units were 12 l/hr, providing approximately 2 hours of detention time in each unit (30 minutes in each stage). The different organic loading rates applied to each unit were controlled by diluting the wastewater with tap water. Influent streams to the four RBC units contained 25 to 100 percent wastewater. In each phase of the study, each unit was operated at constant temperature, influent flow rate, and influent dilution rate. Under these conditions, the process approached steady-state conditions within 3 to 4 weeks.

In each phase of the study after a week of operation, a thin layer of growth covered the discs in the first stages. Generally, in the second week some biomass sloughing was observed in the first stages, and within a few days a new biofilm was built up. After 3 to 4 weeks of operation, the discs in the first stages were covered with a thick, dark-brown, gray biofilm, and further detectable changes in its appearance were not observed. The structure of the biofilm in the first stages and sometimes in the second stages of units C and D (high loading rates) seemed to be spongy rather than a smooth structure. A filamentous growth in these stages may have been the reason for this type of structure.

In the successive stages, the discs were covered with a thinner biofilm layer and were relatively smooth in appearance. In the experiments at temperatures of 15°C and 20°C, the color of the biomass was tan-brown. In the experiment at 5°C, the biomass in the second through the fourth stages had a black-brown appearance. The tan color observed at 15°C and 20°C is probably due to the growth of nitrifiers in these stages. The last stages generally developed a patchy growth. Unlike the attached biomass in the first stages, the attached biomass in the last stages revealed instability with frequent detectable sloughing. At higher temperatures and in the units receiving low organic loadings, instability was stimulated.

The period following stabilization of the attached biomass in the first stage of the four units was considered to represent steady-state conditions. Operation at steady state was continued for 2 to 4 weeks to provide four to six sampling periods. Mechanical failure also resulted in less steady-state data being collected in some stages or units than was planned. In the first phase, motor failure after 7 weeks of operation terminated the steady-state sampling period earlier than was scheduled. Samples from some stages were occasionally discarded because of composite sampler failure or other reasons. The low percentage of wastewater in the influent resulted in low concentrations of filtered chemical oxygen demand (COD) in the effluents. In the units receiving diluted wastewater, the effluent filtered COD values of 15 to 30 mg/l made the chemical analysis very sensitive to interferences such as the presence of nitrites or analytical errors. Regardless of these difficulties, the mean steady-state data for each stage generally consisted of four to six determinations and a low coefficient of variation within these determinations. High variations in the particulate material were observed mainly in the last stages, and the variation was attributed to the unstable attached growth in these stages rather than sample size or analytical errors.

Mode of Operation

Prior to beginning the studies, the hydraulic characteristics of a single stage of one of the RBC units were determined. The test was performed under the same conditions the units were designed to operate during the experiments. Sodium chloride was used as a tracer. A detailed description of the test is presented in Appendix A. The results of this test, shown in Figure A-1 in Appendix A, showed that each single stage of the experimental RBC units can be considered as a complete mix reactor.

As stated earlier, each phase of the study lasted about 2 months with 2 to 4 weeks of operation at steady-state conditions. A summary of the schedule of the project is shown in Table 3.

The influent flow rates were 12 l/hr. About 120 minutes of detention time was provided in the first phase and 140 minutes in the second and third phases. Considering only the projected surface area of the 16 discs in each unit, the hydraulic loading rates were 0.082 $m^3/m^2/d$ (2 gpd/sq ft) in the first phase, and 0.076 $m^3/m^2/d$ (1.9 gpd/sq ft) in the last two phases. Considering all of the available surface area, including the inflation of discs and the area of the supporting side discs, the loading rates reduced to 0.052 $m^3/m^2/d$ (1.3 gpd/sq ft) and to 0.049 $m^3/m^2/d$ (1.2 gpd/sq ft), respectively. The mean steady-state influent loading rates and the wastewater percentages are summarized in Table 4. The coefficients of variation for the flow rates and mixing ratios were less than 3 percent (Table 4). Loading rates were uniform during the entire study and ranged between 0.048 $m^3/m^2/d$ and 0.053 $m^3/m^2/d$. The only significant difference between the first phase (15°C) and the other phases was the hydraulic detention times which were different because equipment failure necessitated replacement of the original equipment. The ratio of the volume to surface area was 4.4 l/m^2 for the first phase and 4.8 l/m^2 for the other phases. At the low hydraulic loading rates used in this study, it is doubtful that these differences (less than 10 percent) in the volume to surface area ratio had any significant effect on the performance efficiency.

Antonie (1976) has reported that performance is not improved with volume to surface area ratios greater than those employed in this study.

Daily flow rates and dilution ratios are summarized in Appendix B (Tables B-1, B-2, and B-3). There was a gradual decline in the liquid temperature as the wastewater flowed through the RBC units. Generally, the temperature decrease was 0.4°C-0.5°C per stage. The temperature decline through the stages was probably due to evaporative heat losses. To obtain the desired mean liquid temperature, the wastewater and tap water influents were maintained at a temperature of 1-2°C higher than the desired average treatment temperature. The temperatures in the first stage were 1°C higher than the desired temperature, and the last stages 1°C less. The mean steady-state data for each stage during each phase are summarized in Tables B-4, B-5, and B-6 (Appendix B). Table 5 summarizes the mean temperatures for the four units for each phase of the experiment.

Influent Wastewater Characteristics

The settled domestic wastewater from Hyrum, Utah, used in this study can be considered a typical weak

Table 3. Summary of the project schedule and the period of steady state conditions.

Phase	Temperature °C	Overall Period	Steady State Period			
			Unit A	Unit B	Unit C	Unit D
I	15	10/25/79 - 12/26/79	12/10-12/19	12/12-12/21	12/11-12/25	12/10-12/25
II	20	2/12/80 - 4/13/80	3/26-4/12	3/27-4/11	3/26-4/12	3/20-4/11
III	5	4/25/80 - 7/10/80	4/29-7/7	5/28-7/9	5/29-7/7	5/28-7/6

Table 4. Summary of mean steady state operating conditions.

Temp. °C	Unit	A			B			C			D**		
	Parameter	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
5	Flow rate ℓ/d	17	286.7	4.6	19	284.2	4.4	17	294.0	5.2	17	302.8	8.2
	Wastewater %	17	66.2	1.6	19	48.5	1.4	17	79.5	1.1		100	
	Detention time-min		141			142			137			133	
	Hydraulic load* $m^3/m^2/d$ (gpd/sq ft)		0.049(1.2)			0.048(1.2)			0.050(1.2)			0.051(1.3)	
15	Flow rates ℓ/d	8	276.3	7.6	8	285.3	5.4	11	282.0	6.8	12	288.6	7.2
	Wastewater %	8	26.1	0.7	8	49.5	0.9	11	73.3	1.6		100	
	Detention time-min		125			121			123			120	
	Hydraulic load* $m^3/m^2/d$ (gpd/sq ft)		0.050(1.2)			0.052(1.3)			0.051(1.3)			0.053(1.3)	
20	Flow rates ℓ/d	12	280.2	3.7	10	283.7	3.7	12	287.4	6.5	12	292.2	4.8
	Wastewater %	12	49.0	0.6	10	69.0	1.3	12	86.0	1.8		100	
	Detention time-min		144			142			140			138	
	Hydraulic load* $m^3/m^2/d$ (gpd/sq ft)		0.048(1.2)			0.048(1.2)			0.049(1.2)			0.050(1.2)	

*All available surface area considered
**Unit D received undiluted sewage
n = number of determinations
S.D. = standard deviation

Table 5. Summary of the observed mean and range of values for the liquid temperature (°C) in the various stages of the four experimental RBC units.

Phase	First Stage		Second Stage		Third Stage		Fourth Stage		Overall
	Mean	Range	Mean	Range	Mean	Range	Mean	Range	Mean
I	16.3	16.0-16.7	15.4	15.0-15.6	14.7	14.4-15.1	14.4	14.1-14.8	15.2
II	20.8	20.5-21.2	20.3	20.0-20.7	19.7	19.1-20.1	19.3	18.6-19.8	20.0
III	5.9	5.4-6.1	5.1	4.7-5.3	4.5	4.2-4.8	4.1	3.8-4.5	4.9

strength domestic wastewater based upon the chemical oxygen demand (COD) concentration, and medium strength in terms of the nitrogen compounds concentrations. The mean COD concentration was approximately 275 mg/l during the first two phases and 175 mg/l during the third phase. The third phase of this study was conducted during the summer. High infiltration rates to the sewerage system in Hyrum, Utah (wastewater source) during the irrigation season account for the low COD concentration observed during the third phase. During the first and second phases of the study, the wastewater did not contain dissolved oxygen. Dissolved oxygen (DO) concentrations of 2.1 to 3.3 mg/l were measured in the wastewater during the third phase, and these high concentrations are also explained by the high infiltration rates. The mean soluble COD measured as filtered COD was about 30 percent of the total COD in the wastewater. During the entire study, the wastewater had a colloidal appearance without any detectable settleable solids. The mean suspended solids (SS) concentrations ranged between 80 and 90 mg/l with about 90 percent of the solids in the form of volatile suspended solids (VSS). The total Kjeldahl nitrogen (TKN) concentration was approximately 40 mg/l and 50 to 75 percent was in the form of ammonia nitrogen (NH_4-N). Nitrite and nitrate nitrogen concentrations were less than 1 mg/l-N. The mean total organic carbon (TOC) concentration in the wastewater was 72 mg/l.

Table 6 contains the mean characteristics of the wastewater for each of the three phases of the project. The detailed data are summarized in Appendix B (Tables B-7, B-8, and B-9). A further description of the influent wastewater characteristics including the particulate material was prepared by calculating the ratios of the various parameters used to describe the pollutional strength of the wastewater. Paired data presented in Tables B-7, B-8, and B-9 were used to calculate the ratios shown in Table 7.

At temperatures of 5°C and 15°C, the particulate material in the wastewater has similar characteristics. The values obtained at a temperature of 20°C may have differed because of the small number of determinations.

The influent substrate concentrations for each unit were calculated for each day the wastewater was sampled by multiplying the influent wastewater percentage times the concentration of the constituent being considered. Tables B-10, B-11, and B-12 in Appendix B contain summaries of the mean steady-state influent characteristics for each unit. Table 8 contains a summary of the influent COD and ammonia nitrogen (NH_4-N) concentrations and the overall organic loading rates expressed as gram COD/m^2/day.

Process Performance

Mixed liquor pH value and dissolved oxygen concentration

The pH value and the dissolved oxygen (DO) concentrations in the mixed liquor of each stage generally revealed a pattern. When nitrification did not occur at a temperature of 5°C, the mixed liquor pH value increased gradually from stage to stage. In the first RBC stages operating at 5°C, the pH value of the mixed liquor ranged between 8.00 and 8.15, and the pH value increased from 8.15 to 8.35 in the mixed liquor of the fourth stages. At the other temperatures, the same pattern was observed in the stages where nitrification was almost completed in the preceding stages. In the stages where nitrification took place, generally there was a decrease in pH value of about 0.05 to 0.15 units per stage. In all of the experiments at temperatures other than 5°C, the mixed liquor pH values were in the range of 7.8-8.1. Table 9 contains a summary of the mean pH values measured in the first and fourth stages. The data presented in Table 9 also indicate a slight decline in the pH value as the organic loading rate increased.

The mixed liquor DO generally increased from stage to stage (Table 10). In units with higher organic loading rates, the DO concentration was lower than that observed in the units with low organic loading rates. In the experiments conducted at 5°C, the DO concentrations were in the range of 6.7 to 9.9 mg/l. At the higher temperatures, the DO values in the first stages ranged from 1.9 to 2.65 mg/l in unit D and 3.6 to 5.0 mg/l in unit A. The decrease in mixed liquor DO with the temperature was probably due to the combined effects of lower DO saturation values resulting in a decrease in the gas transfer rate and the increase in oxygen demand.

Dilution with tap water also may affect the differences in DO concentrations in addition to the organic loading rates. The tap water contained 6.8 to 6.9 mg/l of DO in the 15°C and 20°C experiments, while the wastewater during these phases was devoid of DO. In the 5°C experiments the tap water contained 10.6 to 11.3 mg/l of DO. There were no significant differences in the pH values for the tap water and the wastewater.

Table 6. Summary of the mean influent wastewater characteristics.

Parameter	Temperature, °C								
	5			15			20		
	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
Total COD, mg/ℓ	19	176.6	38.7	12	265.2	96.8	13	289.2	94.9
Filtered COD, mg/ℓ	14	51.4	19.1	9	87.1	37.2	4	83.2	29.3
SS - mg/ℓ	17	78.0	19.3	9	91.9	33.7	5	84.0	38.1
VSS - mg/ℓ	17	75.9	17.0	9	85.2	26.1	5	78.4	32.2
TKN - mg/ℓ	6	44.1	8.0	2	40.5	3.5	9	43.5	8.2
Ammonia-N, mg/ℓ	6	20.13	2.38	3	29.79	0.91	9	22.28	9.89
Nitrite-N, mg/ℓ	6	0.154	0.071	5	0.010	0.003	9	0.238	0.201
Nitrate-N, mg/ℓ	6	0.88	0.59	5	0.01	0.02	9	0.40	0.28
TOC, mg/ℓ							5	72.0	18.7
pH	3	7.78	0.03	1	7.70		2	7.50	0.28
DO, mg/ℓ	2	2.7	0.9	1	0.0		1	0.0	

n = number of determinations

S.D. = standard deviation

Table 7. Summary of the ratios of the characteristics of the wastewater treated by the experimental RBC units.

Parameter	Temperature, °C								
	5			15			20		
	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
$\frac{\text{Filtered COD}}{\text{Total COD}}$	14	0.298	0.161	9	0.367	0.094	4	0.304	0.041
$\frac{\text{Total COD}}{\text{VSS}}$	17	2.372	0.359	9	2.865	0.878	4	3.991	1.801
$\frac{\text{Particulate COD*}}{\text{VSS}}$	14	1.720	0.503	9	1.795	0.596	4	2.832	1.423
$\frac{\text{Total COD}}{\text{TOC}}$							5	3.943	0.698

n = number of determinations

S.D. = standard deviation

* Total COD - Filtered COD

Table 8. Summary of the influent COD and ammonia nitrogen concentrations for each RBC unit and temperature.

Temperature °C	Parameter	Unit A	Unit B	Unit C	Unit D
5	COD - mg/ℓ	118.4	85.6	142.0	173.3
	NH_4-N - mg/ℓ	13.34	9.69	15.98	20.27
	Organic load (g COD/m^2/d)	5.757	4.126	7.081	8.900
15	COD - mg/ℓ	79.3	144.5	192.6	265.2
	NH_4-N - mg/ℓ	7.70	14.76	22.30	29.79
	Organic load (g COD/m^2/d)	3.984	7.496	9.875	13.916
20	COD - mg/ℓ	145.5	202.3	256.7	281.9
	NH_4-N - mg/ℓ	10.0	13.0	17.5	22.3
	Organic load (g COD/m^2/d)	6.915	9.734	12.513	13.971

Table 9. Summary of mean pH values in first and fourth stages of the RBC units.

Temperature (°C)	5		15		20	
Unit	First Stage	Fourth Stage	First Stage	Fourth Stage	First Stage	Fourth Stage
A	7.97	8.22	7.80	8.00	7.95	8.13
B	8.03	8.23	7.70	7.90	7.98	8.08
C	8.00	8.25	7.70	7.73	7.95	8.05
D	8.03	8.27	7.73	7.68	7.80	7.98

Table 10. Summary of dissolved oxygen concentrations (mg/l) in the first and fourth stages of the RBC units.

Temperature (°C)	5		15		20	
Unit	First Stage	Fourth Stage	First Stage	Fourth Stage	First Stage	Fourth Stage
A	7.6	9.4	5.0	7.8	3.6	6.9
B	8.8	9.6	3.9	7.5	2.9	6.5
C	7.9	8.8	3.6	7.9	2.4	6.5
D	7.4	8.3	2.7	7.1	1.9	5.8

The pH value of the tap water was in the range of 7.5 to 8.0.

Carbonaceous substrate removal

In all three phases of this study, the carbonaceous content of the influent wastewater was measured in terms of COD. In the second phase conducted at 20°C, the carbon content also was measured in terms of total organic carbon (TOC). The mixed liquor suspended biomass was measured in terms of volatile suspended solids (VSS). Tables B-16 through B-27 (Appendix B) contain the individual mixed liquor data for total and filtered COD and VSS.

Tables 11, 12, and 13 contain summaries of the mean steady-state data for the experiments conducted at temperatures of 5°C, 15°C, and 20°C, respectively. The coefficients of variation (C.V.) for the filtered COD in the first stages were approximately 15 percent. For the total COD, the C.V. in the first stages was about 20 percent. In the following stages there was a significant increase in the C.V. to values of 25 to 30 percent for filtered and total COD, respectively. The increase reflects the instability in the later stages. The high C.V. for the filtered COD possibly indicated that lysis of biomass was occurring in the later stages and releasing organic material into solutions. The large fluctuations in the total COD were caused by the frequent sloughing of the attached biomass in the later stages contributing to the particulate COD. Because of the inconsistent pattern of sloughing in all stages in each unit, there was no consistent pattern of COD concentrations in the mixed liquor in any of the stages. In the units receiving low organic loading rates, an inconsistent sloughing pattern generally developed in the second stage. In higher loaded units, inconsistent sloughing generally occurred in the third stage and beyond.

Figures 3, 4, and 5 show the mean steady-state mixed liquor filtered COD concentrations when operating the RBC units at 5°C, 15°C, and 20°C, respectively. An analysis of Figures 3, 4, and 5 shows that for the first stages of the units the removal of filtered COD (influent filtered COD minus stage filtered COD) increased when the influent filtered COD was increased. This observation supports the contention that the removal of filtered COD can be described by a substrate limiting reaction. As the temperature was increased, the removal of the filtered COD increased, even beyond 15°C, which is contrary to the results reported by others (Antonie 1976). In stages two through four there was further removal of filtered

Table 11. Summary of steady-state mixed liquor COD and SS data for the RBC units operating at 5°C.

Unit	Parameter	First Stage			Second Stage			Third Stage			Fourth Stage		
		n	Mean	S.D	n	Mean	S.D	n	Mean	S.D	n	Mean	S.D
B	COD Total-mg/ℓ	7	57.3	12.0	7	74.9	39.1	7	58.2	13.4	7	66.0	19.5
	COD Filtered-mg/ℓ	7	21.9	4.6	7	21.2	8.6	7	17.0	3.4	7	19.8	6.8
	SS - mg/ℓ	7	39.8	16.8	7	47.1	34.3	7	38.9	13.2	6	45.7	23.2
	VSS - mg/ℓ	7	36.3	12.5	7	42.0	24.5	7	36.4	10.2	6	40.3	18.3
A	COD Total-mg/ℓ	7	68.6	10.0	7	82.6	27.9	7	68.1	9.1	7	68.0	9.1
	COD Filtered-mg/ℓ	6	25.7	2.5	6	30.4	9.9	7	23.5	1.9	7	23.9	3.6
	SS - mg/ℓ	7	38.9	7.4	7	42.4	9.1	7	41.1	10.8	7	40.3	8.4
	VSS - mg/ℓ	7	37.4	6.8	7	40.3	8.0	7	39.9	9.7	7	37.7	5.6
C	COD Total-mg/ℓ	8	137.9	29.0	7	115.5	21.5	8	122.4	20.5	7	119.5	20.1
	COD Filtered-mg/ℓ	8	30.8	4.4	8	31.0	2.3	7	28.1	4.3	8	28.1	4.6
	SS - mg/ℓ	8	69.6	38.5	8	60.8	20.8	8	68.8	25.7	8	66.0	20.8
	VSS - mg/ℓ	8	62.6	27.9	8	53.6	16.3	8	60.0	18.1	8	58.4	14.1
D	COD Total-mg/ℓ	8	162.3	29.1	8	153.2	23.4	8	164.5	74.4	6	156.6	35.6
	COD Filtered-mg/ℓ	8	39.5	5.9	8	40.0	9.2	8	35.5	4.0	7	36.1	7.6
	SS - mg/ℓ	8	90.1	19.9	8	76.5	22.1	8	88.8	49.5	7	72.1	8.2
	VSS - mg/ℓ	8	79.8	15.7	8	68.9	19.5	8	76.3	34.9	7	62.9	8.5

n = number of determinations

S.D = standard deviation

Table 12. Summary of steady state mixed liquor COD and SS data for the RBC units operating at 15°C.

Unit	Parameter	First Stage			Second Stage			Third Stage			Fourth Stage		
		n	Mean	S.D	n	Mean	S.D	n	Mean	S.D	n	Mean	S.D
A	COD Total-mg/ℓ	2	41.9	7.9	3	66.8	7.0	2	35.0	5.0	1	27	--
	COD Filtered-mg/ℓ	4	21.0	3.4	2	21.9	4.6	2	20.6	0.1	2	21.3	0.3
	SS - mg/ℓ	4	31.5	13.1	4	41.0	27.6	4	28.5	10.1	4	29.5	20.0
	VSS - mg/ℓ	4	30.0	12.8	4	33.5	18.9	4	28.5	10.1	4	26.0	14.8
B	COD Total-mg/ℓ	2	92.5	17.7	3	46.8	4.7	2	55.8	8.5	2	41.5	0.6
	COD Filtered-mg/ℓ	2	28.3	7.5	3	23.2	2.4	2	24.2	0.1	3	21.5	3.5
	SS - mg/ℓ	4	49.5	26.5	4	37.0	19.6	4	26.8	12.4	4	30.8	13.8
	VSS - mg/ℓ	4	46.0	19.8	4	34.8	15.9	4	26.3	12.5	4	29.5	13.3
C	COD Total-mg/ℓ	5	116.6	40.4	5	95.3	23.8	5	90.0	38.3	4	77.1	41.8
	COD Filtered-mg/ℓ	6	38.3	6.9	5	39.7	15.9	5	33.3	5.2	4	25.4	4.0
	SS - mg/ℓ	6	103.2	56.8	6	69.7	32.0	6	89.7	53.3	6	69.8	37.2
	VSS - mg/ℓ	6	89.8	46.4	6	64.0	27.3	6	76.0	43.8	6	62.3	30.6
D	COD Total-mg/ℓ	6	165.0	43.4	5	125.7	31.2	5	135.3	46.3	4	98.1	42.8
	COD Filtered-mg/ℓ	6	49.7	8.2	6	36.5	5.9	6	49.3	20.5	6	39.9	8.3
	SS - mg/ℓ	7	171.6	61.4	7	136.9	28.1	7	113.6	22.9	7	81.7	22.6
	VSS - mg/ℓ	7	143.4	40.3	7	121.1	20.9	7	99.0	18.2	7	74.6	18.2

n = number of determinations

S.D = standard deviation

COD in the higher loaded units, but the removal rate per stage was much less than in the first stage. There was an inconsistent pattern of filtered COD removal in stages two through four probably attributable to cell lysing. In the last stages of the RBC units, the differences in the effluent filtered COD for each of the four units were much smaller than those observed in the first stages.

At higher temperatures there were no significant differences between the effluent filtered COD for each of the four units. Table 14 contains a summary of the filtered COD removal efficiency for the first and fourth stages. Figure 6 presents the removal efficiency as a function of influent filtered COD load. The effect of temperature and the influent loading rate on this relationship is shown in Figure 6. Figures 7, 8, and 9 show the mean steady-state values obtained for total COD at temperatures of 5°C, 15°C, and 20°C, respectively. A general decline in total COD was observed as the wastewater passed through the stages, but with an irregular pattern of decline as mentioned above.

Removal of total COD increased with temperature in the first stages. Overall percentage removal of total COD at 20°C was less than the removals obtained at 15°C because of high fluctuations in the last stages at 20°C. The inconsistent stabilization of the sloughed biomass in the succeeding stages was associated with unstable attached growth in the last stages of the RBC units operating at 20°C. The quantity of attached biomass at 20°C dropped sharply in the third stage in units C and D, while at 15°C there was a gradual decline in attached biomass and about three to four times more biomass than at 20°C. The attached growth characteristics will be discussed later in this chapter. The total COD concentration in the influent did not appear to affect the removal of total COD as was observed in filtered COD removal. Table 15 contains a summary of the total COD removal efficiencies for the first and fourth stages of the RBC units. Figures 10, 11, and 12 show the steady-state mean concentrations of particulate COD (Total COD - Filtered COD) in the four stages of the RBC units operating at 5°C, 15°C, and 20°C, respectively.

Table 13. Summary of steady state mixed liquor COD and SS data for the RBC units operating at 20°C.

Unit	Parameter	First Stage			Second Stage			Third Stage			Fourth Stage		
		n	Mean	S.D	n	Mean	S.D	n	Mean	S.D	n	Mean	S.D
A	COD Total-mg/ℓ	4	121.3	53.5	5	86.5	29.5	5	92.2	22.1	5	88.6	11.9
	COD Filtered-mg/ℓ	5	25.2	5.8	5	21.6	4.3	5	31.6	15.7	4	25.4	11.5
	SS - mg/ℓ	5	64.4	20.3	5	66	25.5	5	63.4	19.9	5	57.2	15.0
	VSS - mg/ℓ	5	55	13.9	5	56	16.6	5	56.2	14.7	5	51.0	9.6
B	COD Total-mg/ℓ	5	98.0	20.7	4	88.0	34.5	4	73.5	15.6	4	123.2	37.4
	COD Filtered-mg/ℓ	4	26.8	2.9	4	36.3	11.1	4	32.3	9.6	5	27.8	11.6
	SS - mg/ℓ	5	61.8	15.0	5	57.4	14.5	5	58.2	16.4	5	67.4	18.0
	VSS - mg/ℓ	5	58.6	13.8	5	50.6	7.2	5	52.4	12.6	5	61.4	11.8
C	COD Total-mg/ℓ	5	106.3	20.6	5	143.1	42.2	5	108.2	46.4	4	122.2	50.4
	COD Filtered-mg/ℓ	3	38.1	0.8	3	29.6	1.5	4	26.3	3.3	4	27.9	3.4
	SS - mg/ℓ	5	87.6	22.5	5	90.6	33.5	5	82.4	19.2	5	91.2	22.6
	VSS - mg/ℓ	5	78.2	17.6	5	79.0	23.1	5	69.8	13.9	5	78.2	16.7
D	COD Total-mg/ℓ	6	194.5	61.7	5	171.2	43.0	5	121.2	30.3	5	187.5	56.3
	COD Filtered-mg/ℓ	6	40.4	4.0	5	28.1	3.0	4	31.6	6.1	5	28.0	5.9
	SS - mg/ℓ	6	116.2	73.8	6	126.2	66.2	6	101.7	49.1	6	115.7	34.6
	VSS - mg/ℓ	6	97.8	53.5	6	106	49.8	6	89.0	38.7	6	98.3	25.2

n = number of determinations

S.D = standard deviation

Particulate COD removal occurred as the wastewater flowed through the stages, and the pattern of removal was similar to that observed for the total COD. Figures 13, 14, and 15 show the mean steady-state concentrations of VSS in the mixed liquor when operating the RBC units at temperatures of 5°C, 15°C, and 20°C, respectively. In general, in the first stage there was an increase in the VSS concentration indicating sloughing of attached biomass. There were significant differences in the characteristics of the VSS at 20°C and at 15°C. At 20°C there were less variations in the VSS concentrations between the stages than was observed at 15°C. This may be attributable to differences between yield coefficients and biomass stabilization factors as will be discussed in a later section. Although it appears that there were no significant changes in the mixed liquor VSS concentrations as the wastewater flowed through the stages, the particulate COD decreased indicating that the sloughed biomass was being stabilized. This same phenomenon has been observed in a full scale system (Malhotra et al. 1975).

Although the mixed liquor contained sloughed biomass, the removal of the influent particulate COD in the first stage implies that the influent particulate COD as well as the soluble COD is available substrate. In addition, the major part of the carbon in the influent wastewater was in colloidal form. To incorporate the influent total COD in an analysis of the process requires measurements of the unconsumed particulate substrate. Unfortunately, this kind of measurement is difficult to perform. Traditionally, the approach is to assume that the particulate material in the mixed liquor is biomass and the remaining substrate is the soluble substrate.

During the entire study, the mixed liquor SS were significantly different from the wastewater SS. The mixed liquor SS were in the form of flocs, which tended to settle within a few minutes rather than the colloidal form of the wastewater SS. Also, the ratio of the particulate COD to VSS in the mixed liquor was different from the ratio in the wastewater. For 5°C and 15°C these differences are statistically significant at levels of 15 and 2 percent, respectively. As discussed earlier, the ratio of the particulate COD to VSS obtained at 20°C was unreliable, and t-tests were not performed. Table 16 shows the mean ratios of the particulate COD to VSS in the first stages of the RBC

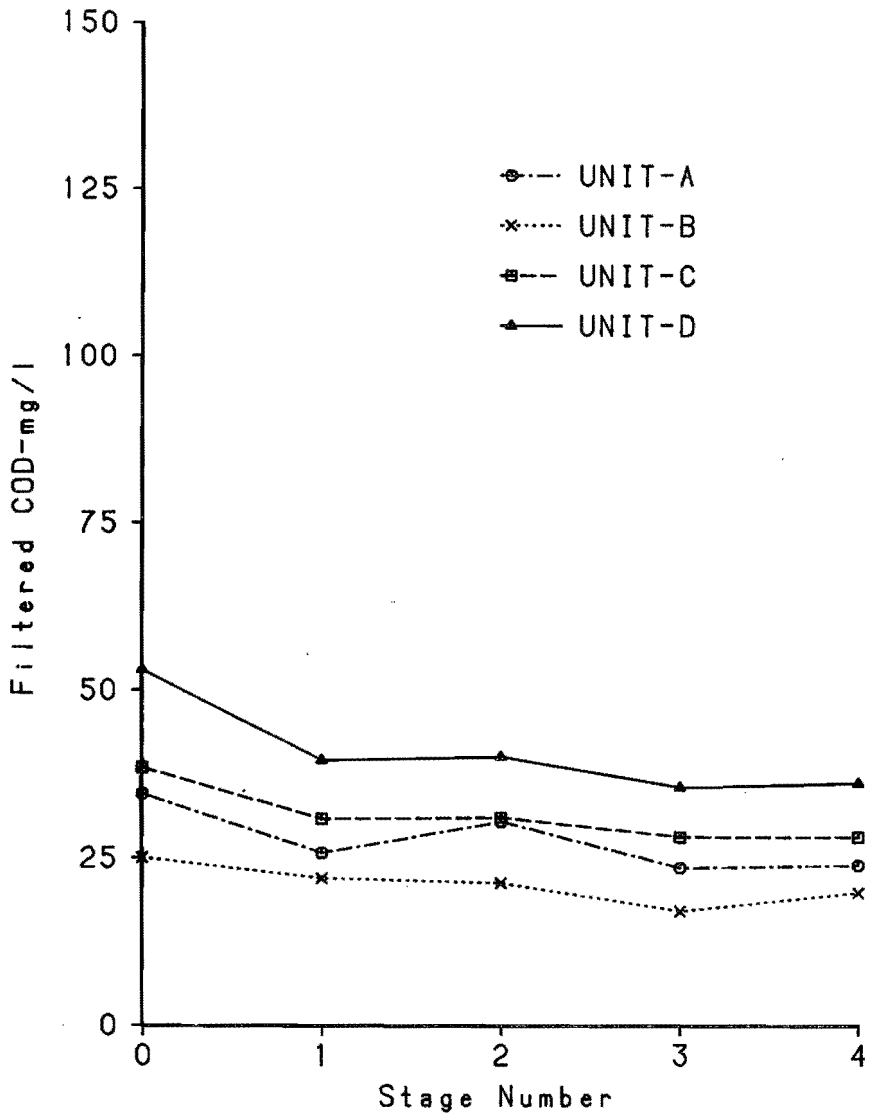


Figure 3. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 5°C.

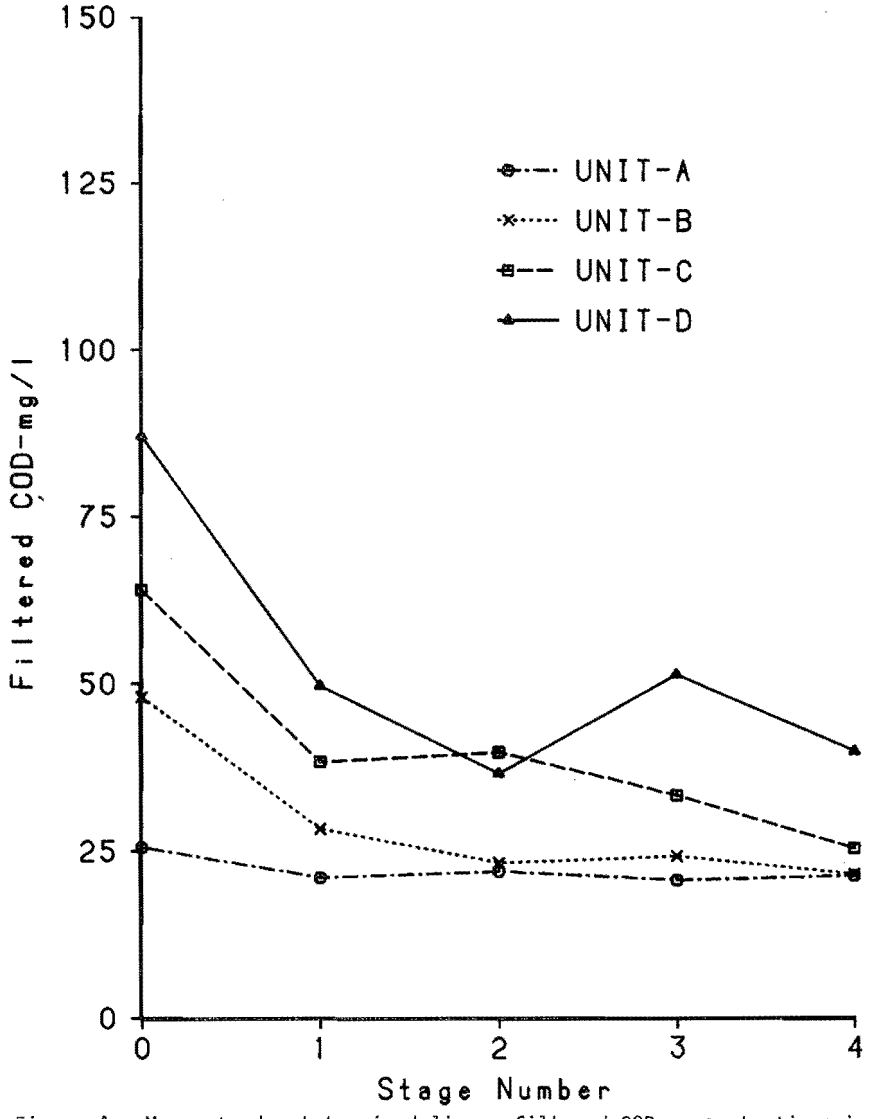


Figure 4. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 15°C.

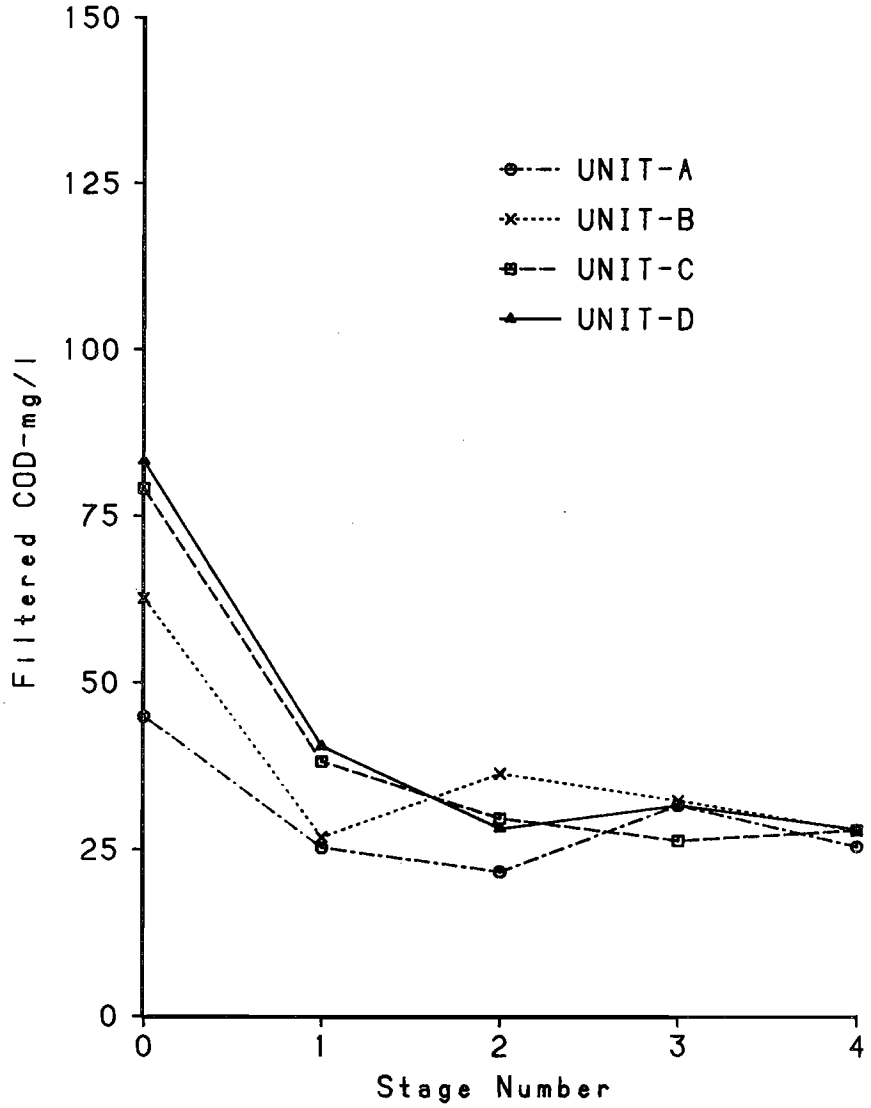


Figure 5. Mean steady-state mixed liquor filtered COD concentrations in the four stages of the RBC units operating at 20°C.

Table 14. Summary of filtered COD removal efficiency (%) at steady-state for each RBC unit, stage, and temperature.

Unit	Temperature °C	5		15			20		
	Filtered COD Influent load	Removal		Filtered COD Influent load	Removal		Filtered COD Influent load	Removal	
	g/d/m²	1st stage	Final	g/d/m²	1st stage	Final	g/d/m²	1st stage	Final
A	1.678	25.5	30.7	1.281	17.7	16.5	2.134	43.9	43.4
B	1.205	12.4	20.8	2.490	41.0	55.2	3.012	57.2	55.6
C	1.915	19.8	26.8	3.282	40.2	60.3	3.856	51.8	64.7
D	2.722	25.5	31.9	4.570	42.9	54.2	4.123	51.4	66.4

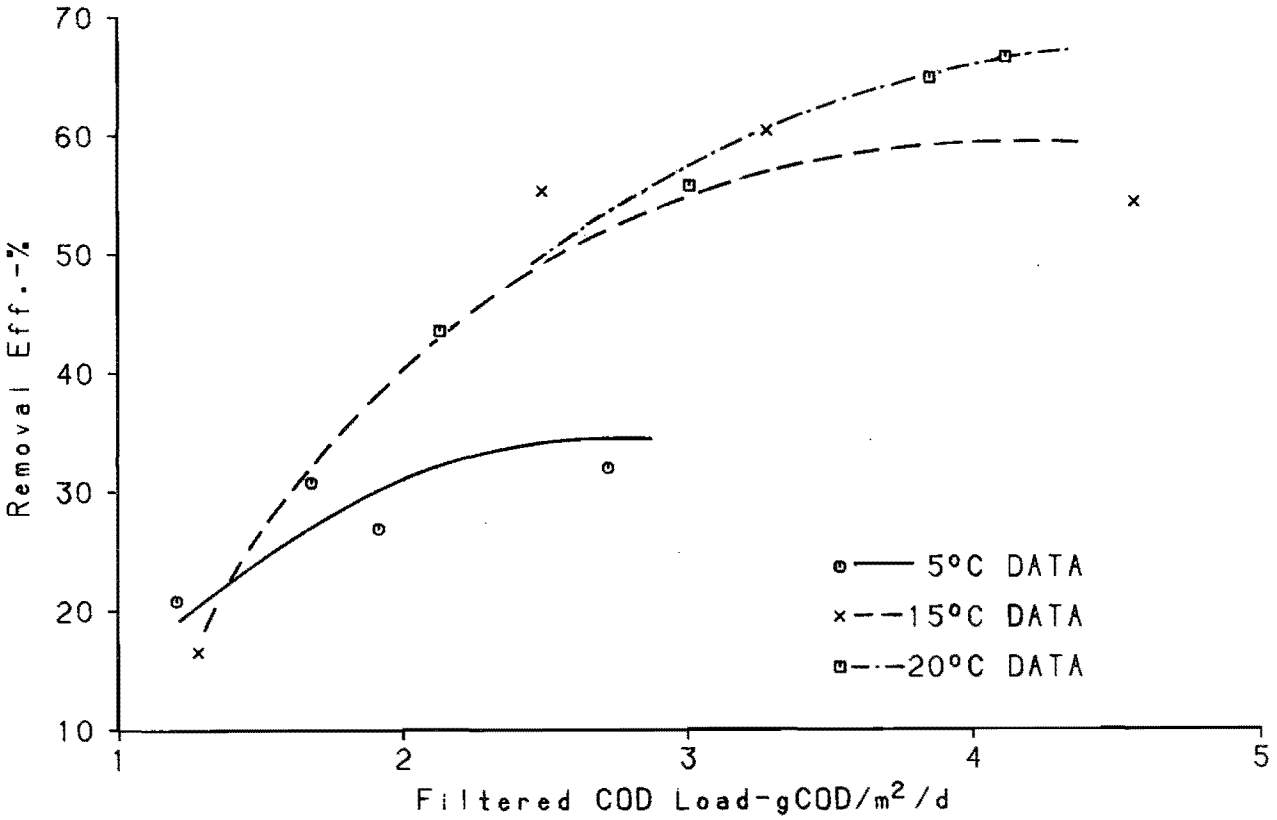


Figure 6. The effect of temperature and organic loading rate on final filtered COD removal efficiency.

Table 15. Summary of total COD removal efficiency (%) at steady state for each RBC unit, stage and temperature.

Temperature Unit	5°C		15°C		20°C	
	First Stage	Final	First Stage	Final	First Stage	Final
A	42.1	42.6	47.2	66.0	34.3	39.1
B	33.1	22.9	36.0	71.3	51.6	39.1
C	2.9	15.9	39.5	66.0	58.6	52.4
D	6.4	9.6	37.8	63.0	31.0	33.5

Table 16. The mean ratios of the particulate COD to VSS in the mixed liquor of the first stages of the RBC units.

Temperature °C	5	15	20
Number of determinations	29	13	16
Mean mg/l COD/mg/l VSS	1.434	1.079	1.366
Standard deviation	0.554	0.613	0.654

units. Based upon the characteristics of the VSS produced in the RBC units, it appears justifiable to consider the mixed liquor VSS as sloughed biomass and the influent total COD as the available substrate. Table 17 shows the substrate removal efficiencies when considering total COD as the available influent substrate and the filtered COD as the remaining substrate in the effluent from the RBC units.

Figure 16 shows the substrate removal efficiencies as a function of influent total organic loadings. The relationship supports the contention that a saturation type relationship can be used to describe the substrate limiting conditions. Again, the removal efficiency increased as the temperature increased, including temperatures beyond 15°C. The effect of temperature between 15°C and 20°C was observed only at organic loadings greater than 10 gCOD/d/m^2 when considering the overall performance.

Figure 17 shows the relationship between filtered COD concentrations in the effluent and the organic loading rate and temperature. The same effects mentioned before are evident in Figure 17. A linear relationship exists between the overall removal of COD in terms of grams of COD removed per unit area and the influent substrate loading rate (Table 18). The slopes of the relationships increased as the temperature increased showing the temperature dependence of the substrate removal performance. The slopes differ at a significance level of 1 percent. A similar relationship was obtained with a full scale RBC plant

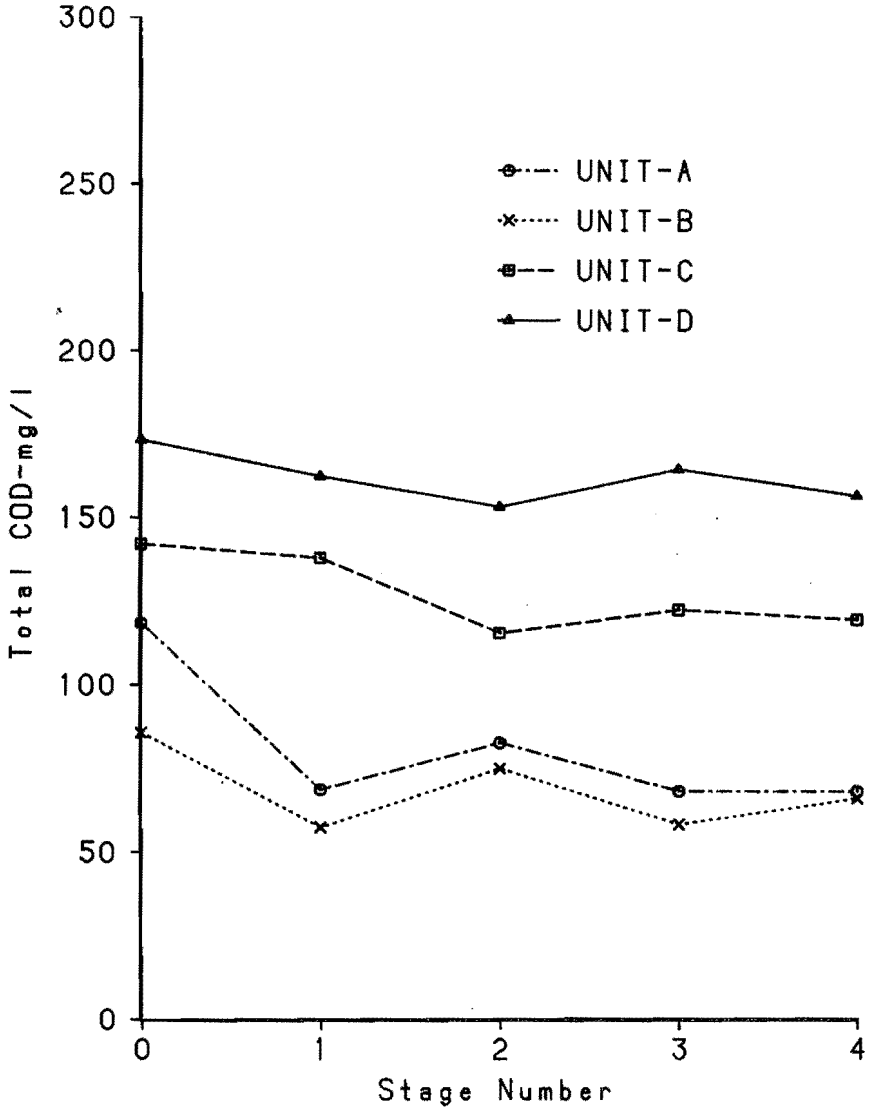


Figure 7. Mean steady-state mixed liquor total COD concentrations in the four stages of the RBC units operating at 5°C.

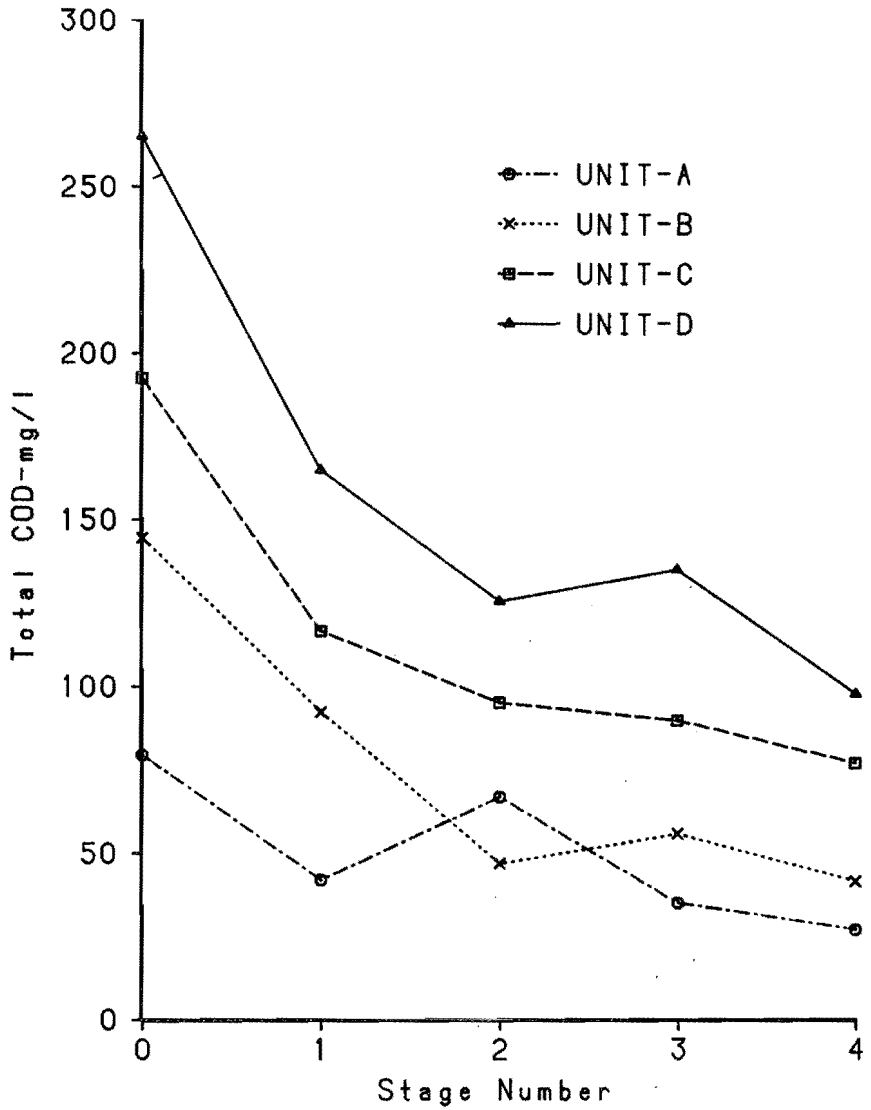


Figure 8. Mean steady-state mixed liquor total COD concentrations in the four stages of the RBC units operating at 15°C.

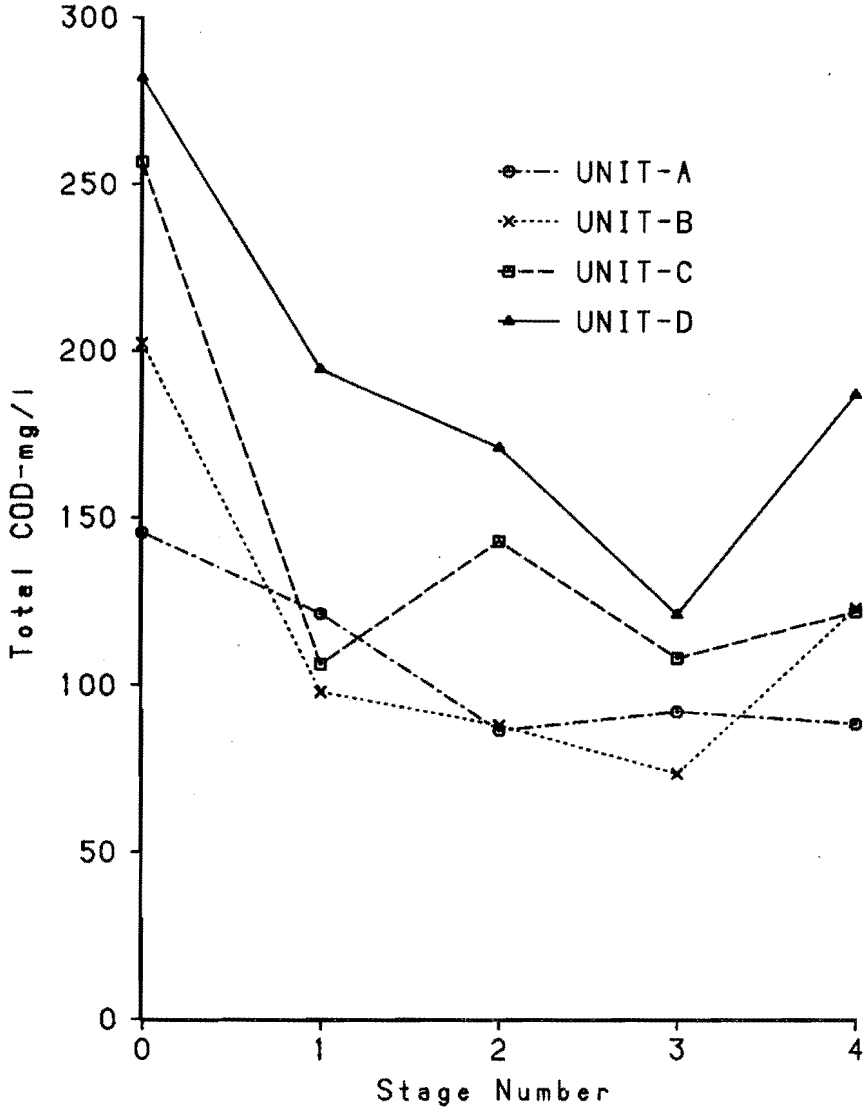


Figure 9. Mean steady-state mixed liquor total COD concentrations in the four stages of the RBC units operating at 20°C.

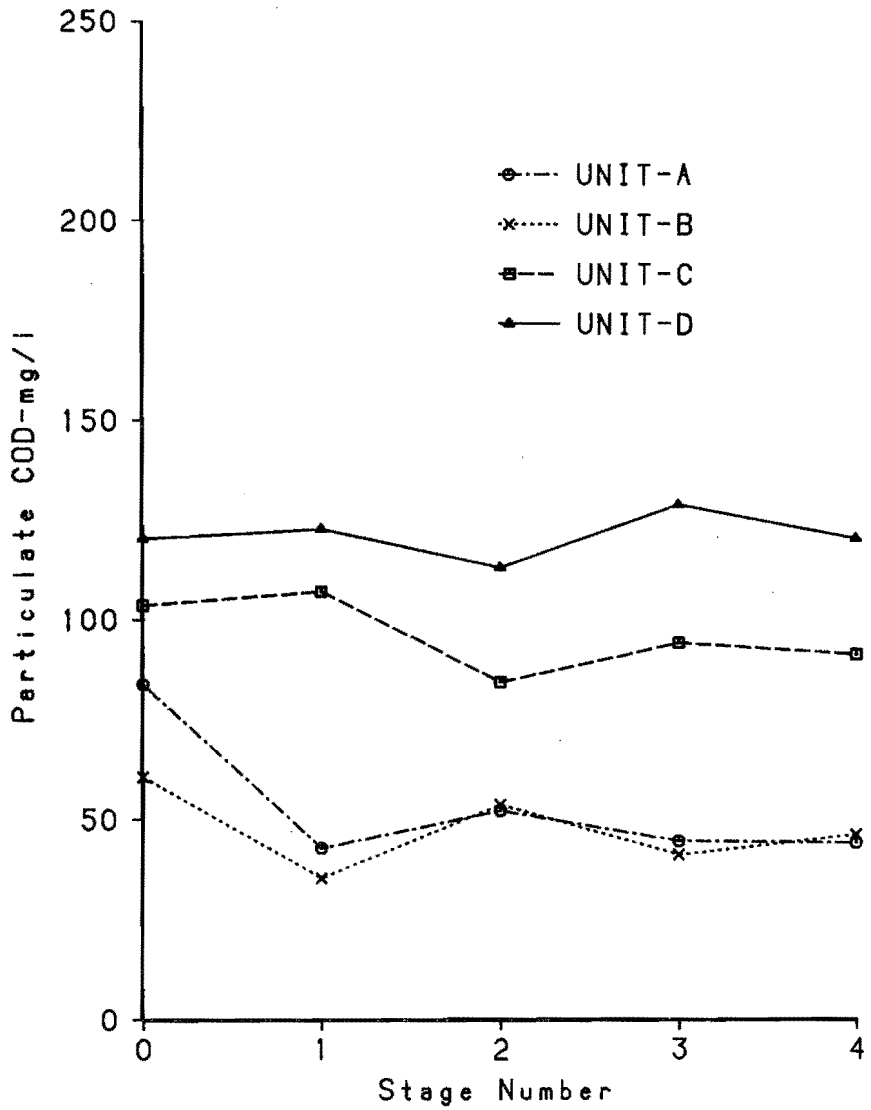


Figure 10. Mean steady-state mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 5°C.

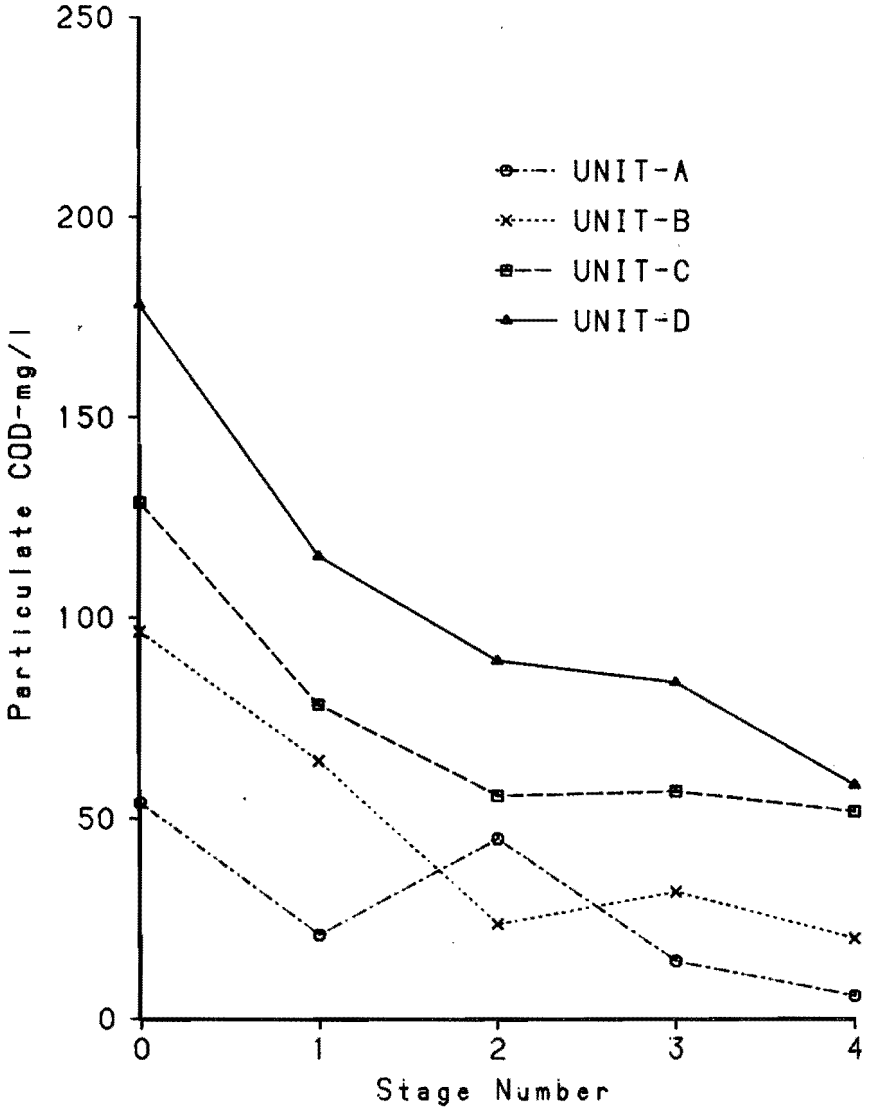


Figure 11. Mean steady-state mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 5°C.

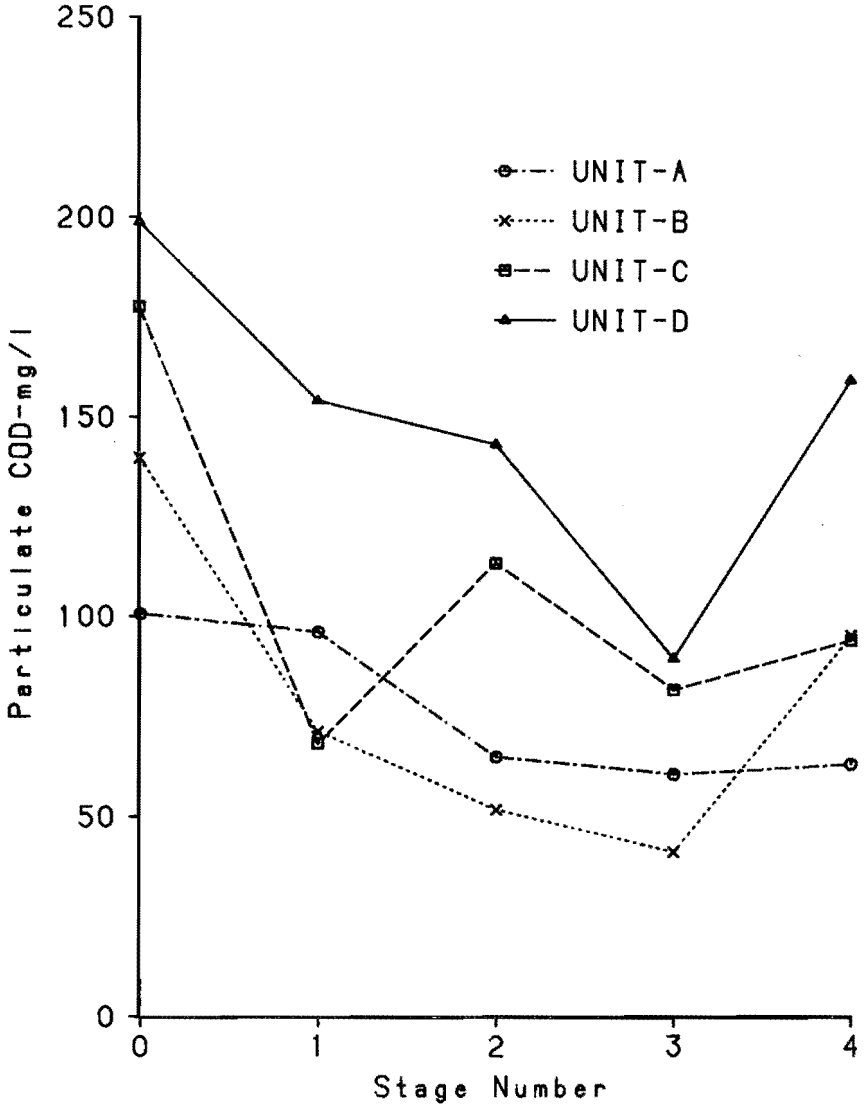


Figure 12. Mean steady-state mixed liquor particulate COD concentrations in the four stages of the RBC units operating at 20°C.

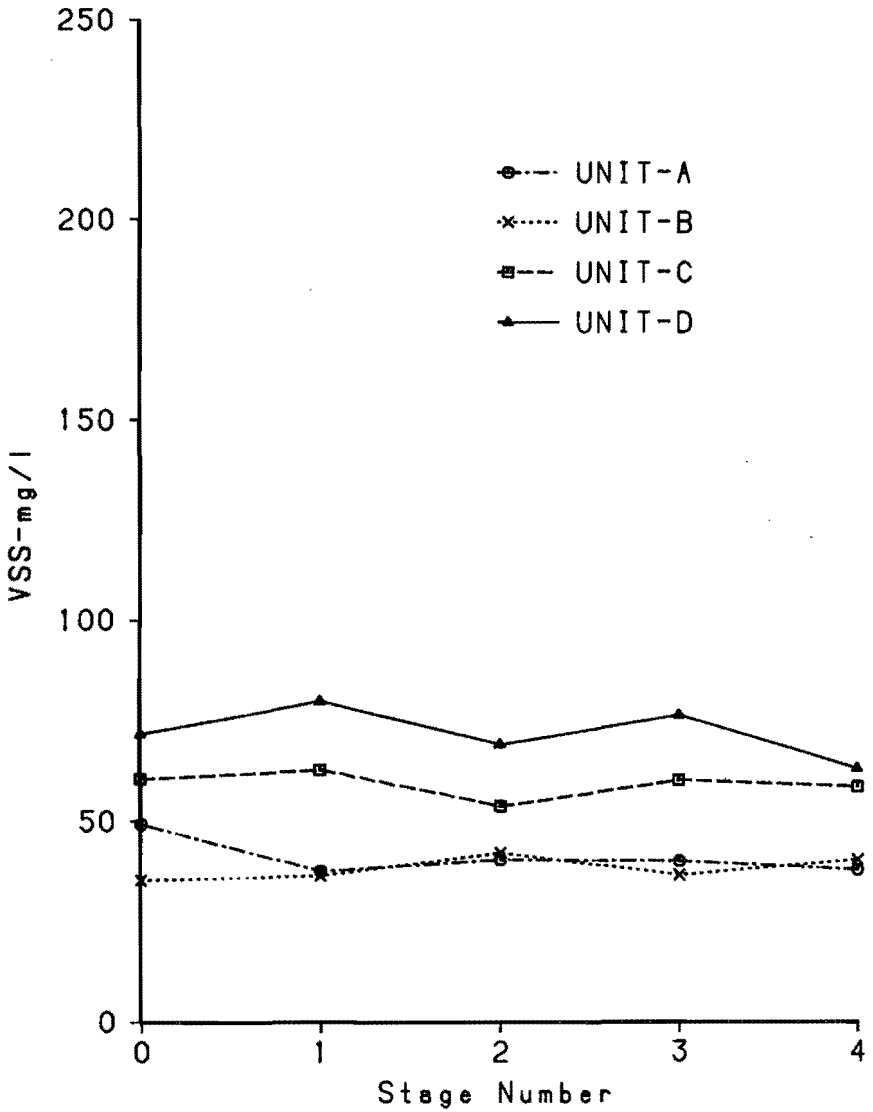


Figure 13. Mean steady-state mixed liquor VSS concentrations in the four stages of the RBC units operating at 5°C.

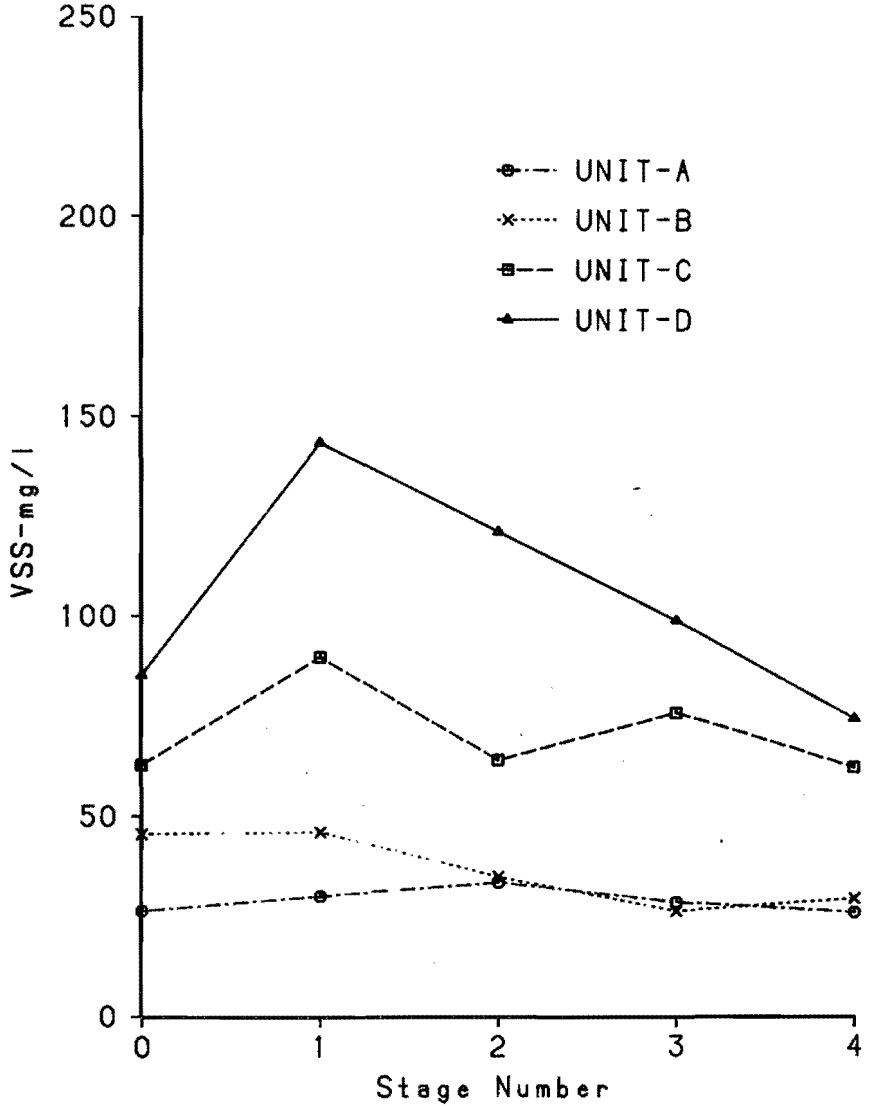


Figure 14. Mean steady-state mixed liquor VSS concentrations in the four stages of the RBC units operating at 15°C.

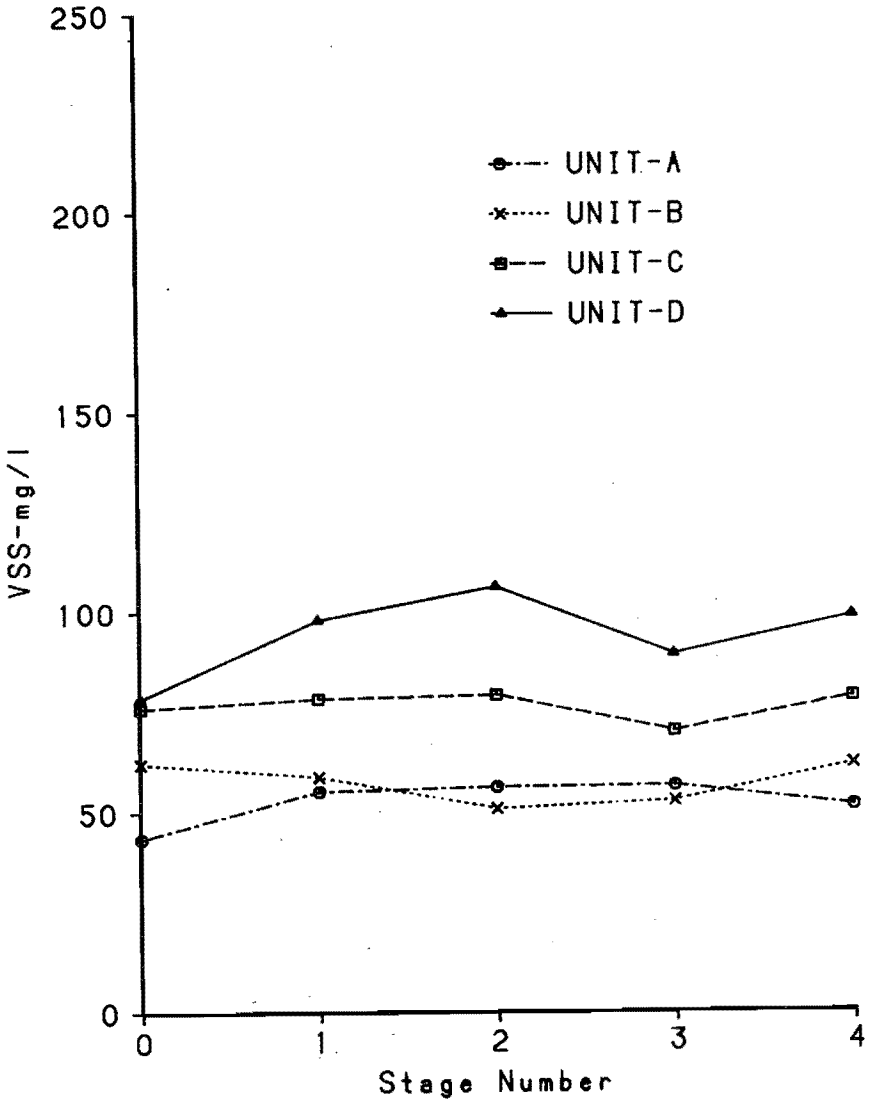


Figure 15. Mean steady-state mixed liquor VSS concentrations in the four stages of the RBC units operating at 20°C.

Table 17. Summary of COD removal efficiencies.

Temperature Unit	5°C		15°C		20°C	
	First Stage	Overall	First Stage	Overall	First Stage	Overall
A	78.3	79.8	73.5	73.1	82.7	82.5
B	74.4	76.9	80.4	85.1	86.8	86.3
C	78.3	80.2	80.1	86.8	85.2	89.1
D	77.2	79.2	81.3	85.0	85.7	90.0

Table 18. Correlation of COD removal per unit area with influent loading rate.

Temperature °C	Unit	Influent Load gCOD/m²/d	Overall Removal gCOD/m²/d	Linear Regression Analysis	
				R^2	Slope
5	B	4.126	3.172	0.999	0.8113
	A	5.757	4.595		
	C	7.081	5.680		
	D	8.900	7.046		
15	A	3.984	2.914	0.997	0.8965
	B	7.496	6.380		
	C	9.875	8.573		
	D	13.916	11.822		
20	A	6.915	5.708	0.9999	0.9759
	B	9.734	8.397		
	C	12.513	11.153		
	D	13.971	12.583		

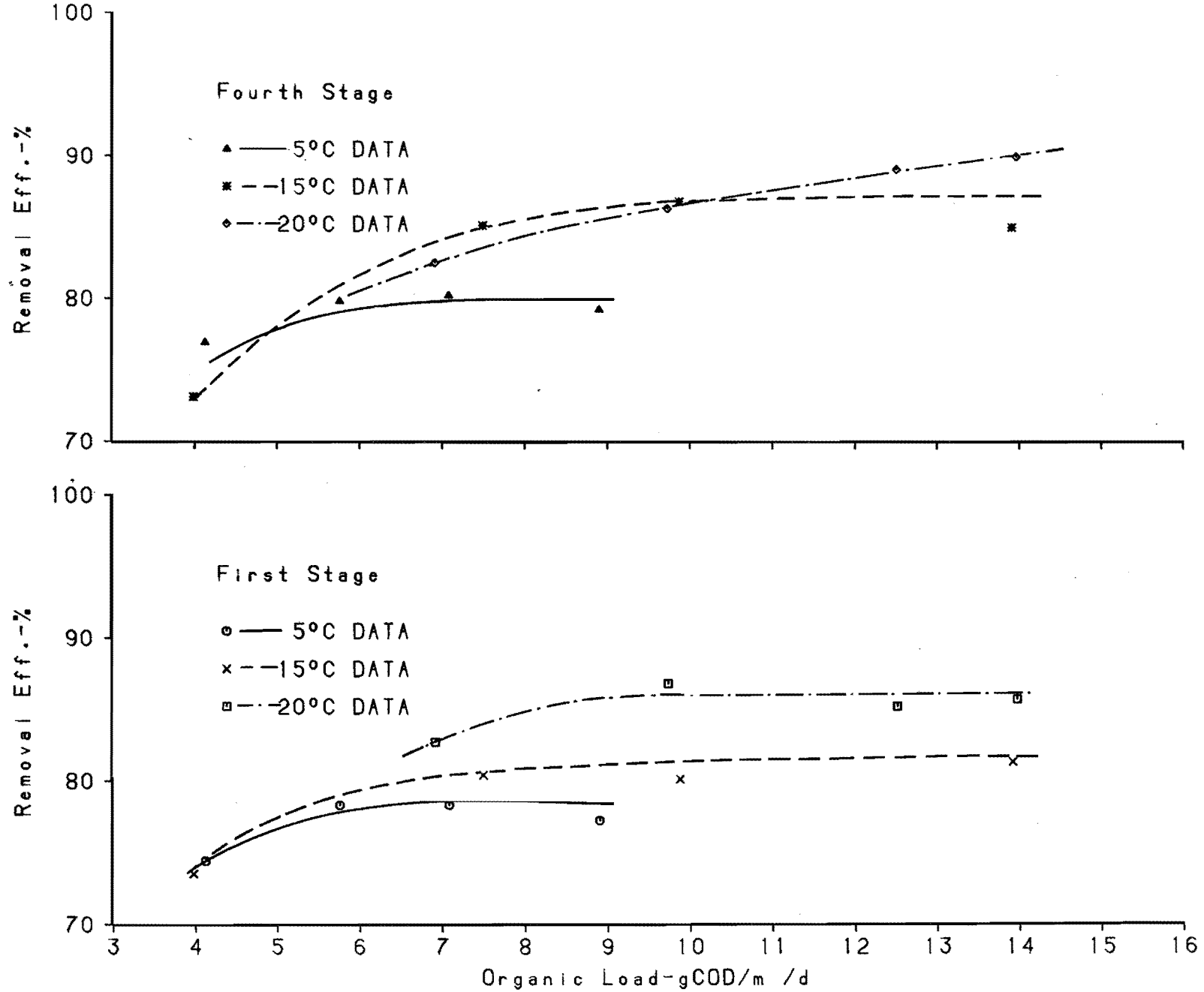


Figure 16. The effect of temperature and organic loading rate on the carbonaceous substrate removal (as total COD) in the first and fourth stages of the RBC units.

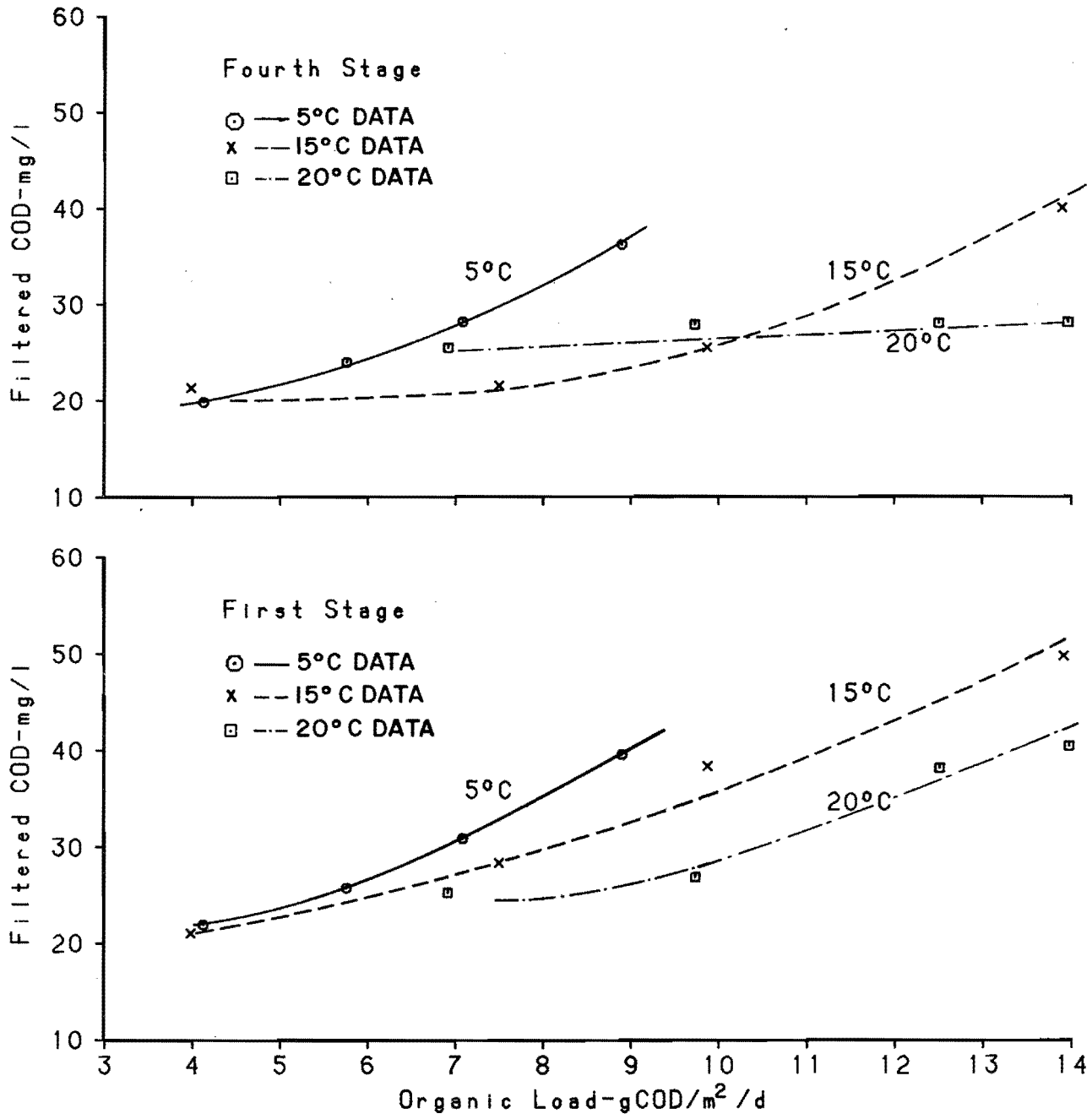


Figure 17. The effect of temperature and organic loading rate on the removal of filtered COD in the first and fourth stages of the RBC units.

treating municipal wastewater at Kirksville, Missouri (Dupont and McKinney 1980). The substrate concentration was measured as BOD_5, and the slope of the relationship was 0.893 dimensionless based upon data collected over 2 years. Figure 18 shows the relationship between substrate removal rate and organic loading rate. All three relationships are significant at the 1 percent level.

An important factor in the engineering application of the RBC process is the apparent sludge production. Table 19 contains a summary of the mass of sludge produced per mass of substrate removed in terms of grams SS/g COD removed. The mean sludge production was 0.55 for 5°C, 0.430 (excluding unit B) for 15°C, and 0.429 for 20°C. The increase in sludge production at lower temperatures was probably due to lower decay rates. Yield coefficients and decay rates are discussed in a later section. Antonie (1976) related sludge production to BOD_5 removal, reported values of 0.5 to 0.6 grams of SS/g of BOD_5 removed, and observed that the sludge production increased at lower temperatures. Although BOD_5 and COD relationships are not directly comparable, the results indi-

Table 19. Summary of sludge production per unit of influent substrate removed (gSS/gCOD).

Temperature Unit	5°C	15°C	20°C
A	0.427	0.509	0.476
B	0.695	0.250	0.386
C	0.580	0.418	0.399
D	0.526	0.363	0.456

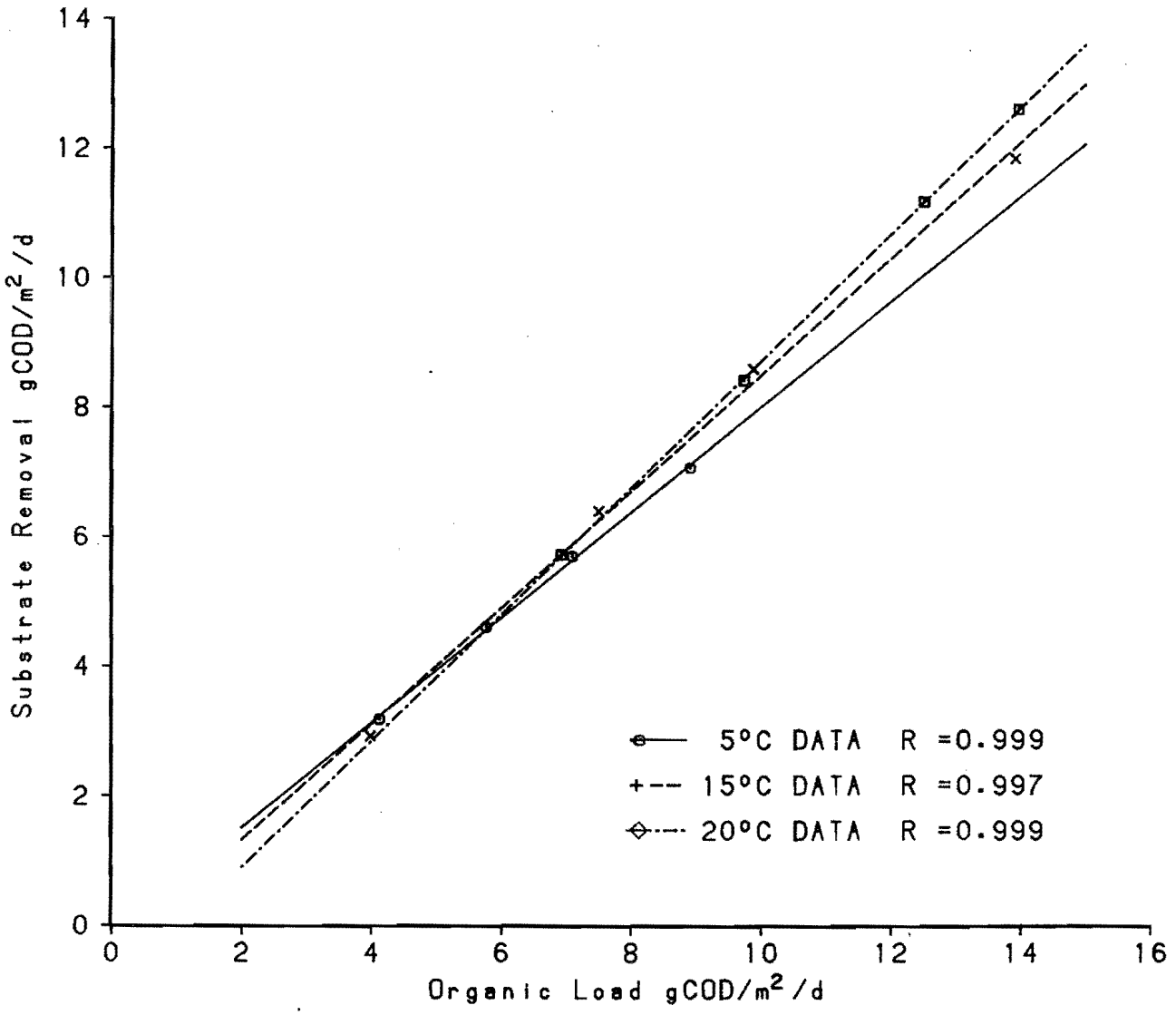


Figure 18. Overall substrate removal versus organic loading rate.

cate reasonable agreement between the experimental results and the work by Antonie (1976).

At 20°C, during the period of 3/19 to 4/8, samples from the experimental systems also were analyzed for TOC. The individual data are summarized in Tables B-28 and B-29 (Appendix B). A summary of the mean TOC data is presented in Table 20 and plotted in Figures 19 and 20. The irregular pattern observed for the total organic carbon concentrations in the second through the fourth stages is shown in Figure 19. Similar irregular patterns were observed for the COD measurements. The filtered organic carbon concentrations were not reduced after the first two stages (Figure 20). The same results were obtained with the filtered COD. Table 21 contains a summary of the TOC removal efficiencies. The major part of the TOC removal occurred in the first stages. The overall TOC removal ranged from 42 percent to 64 percent, and the substrate organic carbon removal (substrate organic carbon removal = influent total organic carbon - effluent filtered organic carbon) ranged from 73 percent to 81 percent. In the first stages the TOC removal appeared to follow a saturation type substrate limiting phenomenon, but in the latter stages such a trend was not observed.

Table 20. Summary of the steady-state mean TOC data for each stage of the RBC units operating at 20°C.

Unit	Total Organic Carbon-mg/l			Filtered Organic Carbon-mg/l		
	n	Mean	S.D.	n	Mean	S.D.
A						
Influent	6	34.0	8.7	3	14.5	2.5
First Stage	3	29.2	15.6	3	14.9	1.6
Second Stage	2	10.7	1.2	3	9.4	1.4
Third Stage	2	11.8	1.9	2	8.3	0.5
Fourth Stage	2	16.9	0.1	2	8.4	0.2
B						
Influent	4	45.3	8.5	2	19.4	2.9
First Stage	2	22.4	2.6	3	14.1	1.7
Second Stage	2	19.6	6.6	3	12.5	2.7
Third Stage	2	18.6	5.7	2	11.9	4.7
Fourth Stage	2	24.4	10.7	2	11.4	4.8
C						
Influent	6	60.2	15.8	3	25.5	4.3
First Stage	3	31.1	3.6	3	17.5	0.5
Second Stage	2	24.1	1.2	3	16.9	3.5
Third Stage	2	38.4	6.2	2	17.7	5.4
Fourth Stage	2	21.6	1.1	2	16.2	4.8
D						
Influent	4	65.6	13.9	2	27.6	4.9
First Stage	3	29.8	5.7	3	20.7	2.7
Second Stage	2	39.9	1.8	3	14.9	2.3
Third Stage	2	30.2	5.2	2	14.0	0.9
Fourth Stage	2	37.9	2.6	2	12.3	0.2

n = number of determinations

S.D. = standard deviation

Table 21. Summary of TOC removal efficiency (%) at steady-state for each RBC unit operating at 20°C.

Unit	Filtered Organic Carbon		Total Organic Carbon		Substrate Organic Carbon[a]	
	First Stage	Overall	First Stage	Overall	First Stage	Overall
A	<0	57.9	14.1	50.3	56.2	75.3
B	27.3	41.2	50.6	46.1	68.9	74.8
C	31.4	36.5	48.3	64.1	70.9	73.1
D	25.0	55.4	54.6	42.2	68.5	81.3

At 20°C the samples from the frist two stages were analyzed for filtered carbonaceous biological oxygen demand (BOD_5) and the results are shown in Table 22 along with the corresponding filtered COD values. Figure 21 shows the relationship between these two measurements that were used as indices of the carbonaceous content of the substrate. The low values of filtered BOD_5 (5 - 7 mg/l) measured in the first stages of units A, B, and C were not further reduced in the second stages. The same results were obtained with the filtered COD and filtered TOC.

The similar low filtered BOD_5 concentrations obtained in the second stages of the four units emphasized that there were no significant differences in substrate concentrations beyond the first stages. The nonlinear curve in Figure 21 shows a decline in the relationship between filtered BOD_5 and filtered COD as the filtered COD decreased. An extrapolation of the curve passes through the origin, indicating that even low filtered COD values in the first two stages could be used as an indicator of carbonaceous substrate for domestic wastes. An exception occurred in the second stage of unit B and was probably due to lysis of cell material and the release of nonbiodegradable material into solution.

Ammonia nitrogen removal

During the first phase of the study conducted at 15°C, it was observed that significant quantities of

Table 22. Summary of the steady-state filtered BOD_5 concentrations in the first and second stages of the RBC units operating at 20°C.

Unit-Stage	Date	Filtered BOD_5 mg/l	Filtered COD mg/l
A-1	4/8	5.0	24.0
A-2	4/8	4.0	19.9
B-1	4/7	7.0	30.1
B-2	4/7	4.0	42.2
C-1	4/8	7.0	37.3
C-2	4/8	5.0	30.7
D-1	3/20	8.0	35.2
D-2	3/20	3.0	25.6
D-1	4/7	13.0	46.5
D-2	4/7	4.0	26.2

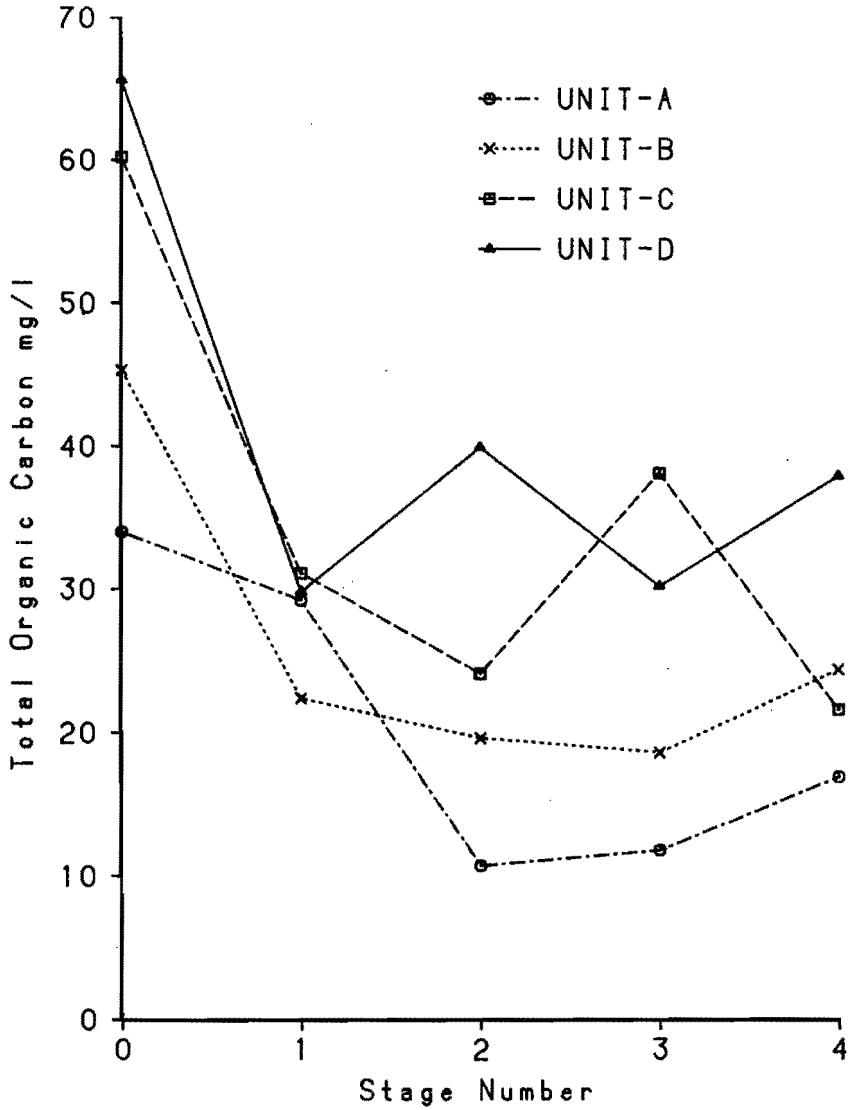


Figure 19. Mean steady-state mixed liquor TOC concentrations in the four stages of the RBC units operating at 20°C.

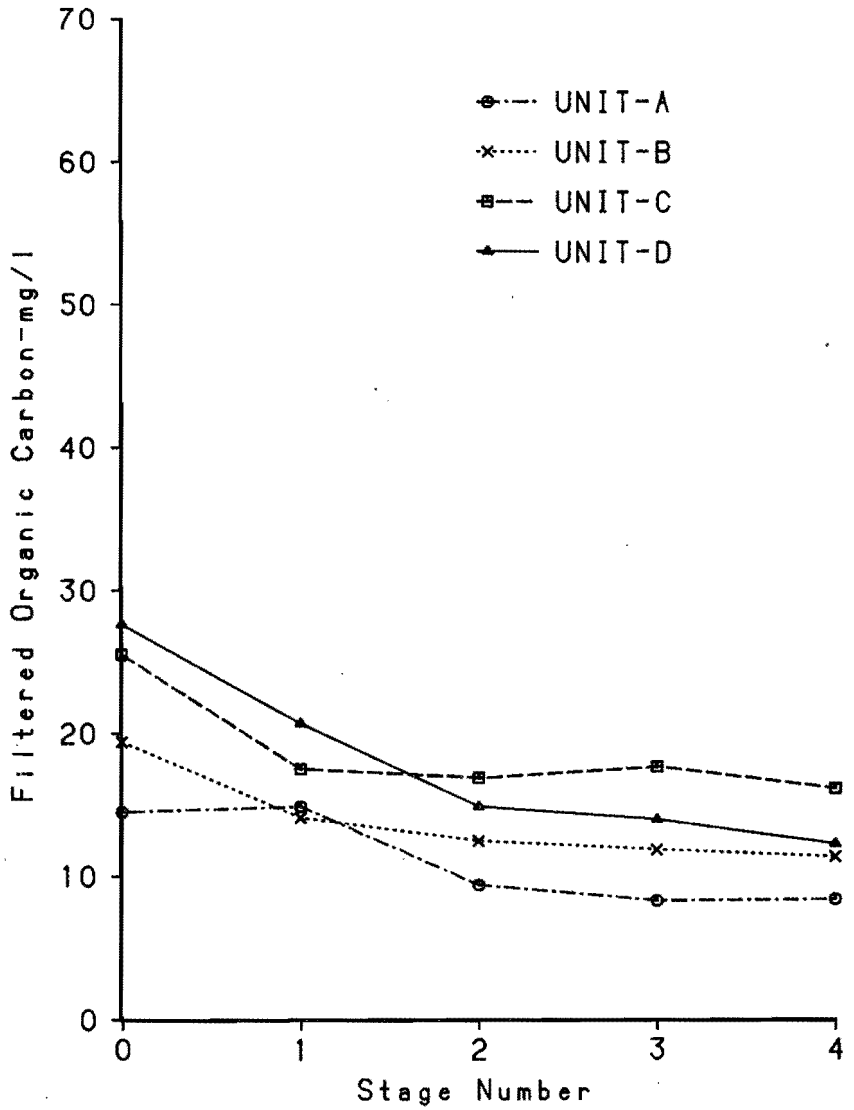


Figure 20. Mean steady-state mixed liquor filtered organic carbon concentrations in the four stages of the RBC units operating at 20°C.

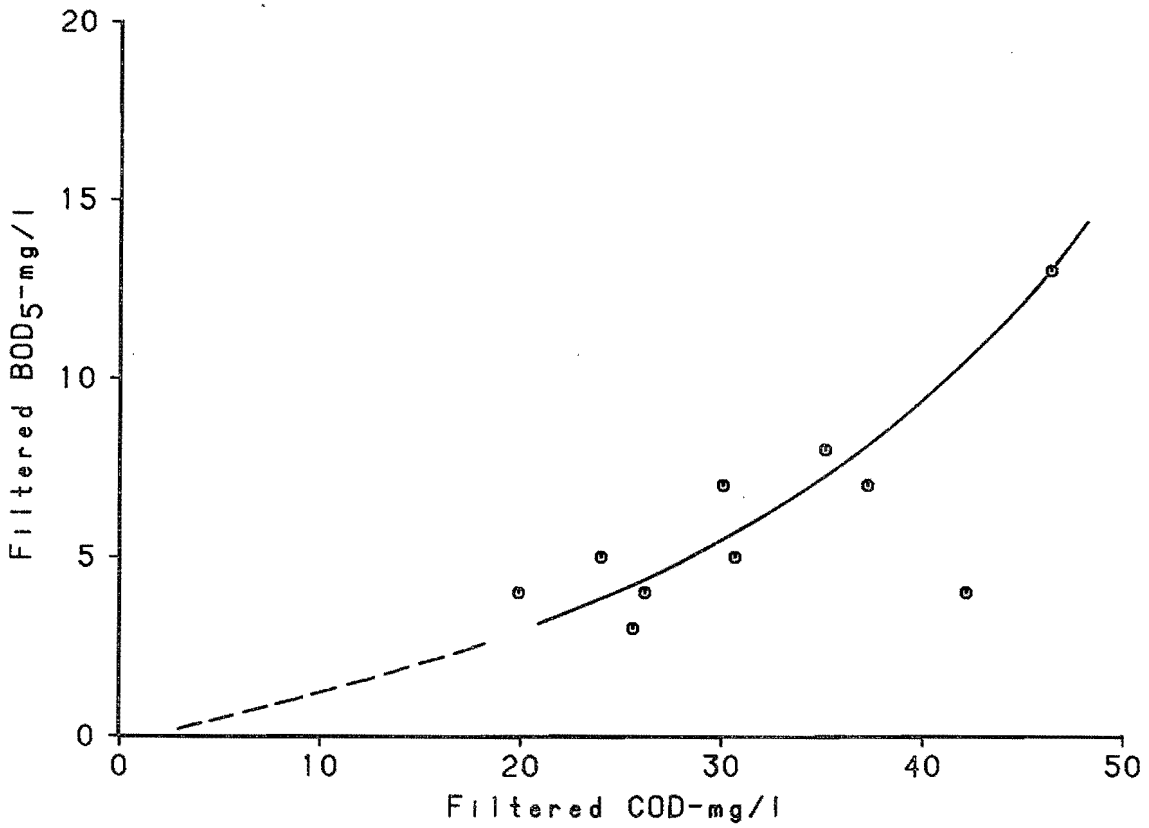


Figure 21. The relationship between filtered BOD_5 and filtered COD in the first two stages of the RBC units operating at 20°C.

ammonia nitrogen were being converted to other forms or being removed in the experimental RBC units. Both at 15°C and 20°C, the majority of the ammonia nitrogen was converted to nitrate nitrogen. Other processes that may contribute to ammonia nitrogen removal are ammonia nitrogen assimilation and ammonia gas stripping. However, these processes had little influence in this study as will be seen later. At a temperature of 5°C, there were no significant changes in ammonia nitrogen and in NO_x (sum of nitrite nitrogen and nitrate nitrogen) concentrations between the influent and final effluent. Only in unit B, which received a low organic loading rate, was there slight ammonia nitrogen removal (0.5 mg/l). As the wastewater flowed through the stages of units A, B, and C, there were indications that low rates of nitrification were occurring.

Tables 23, 24, and 25 summarize the steady-state concentrations for the nitrogen compounds in the stages of the four experimental RBC units. In the second phase, conducted at 20°C, the influent ammonia nitrogen concentrations varied widely, and there were two distinct sampling periods with two concentrations. In the first period the influent contained approximately 35 mg/l ammonia nitrogen, and in the second period the concentration was about 15 mg/l. Table B-30 (Appendix B) contains a summary of the influent nitrogen compounds concentrations and Table 25 summarizes the RBC data for the second sampling period when most of the nitrogen data were collected.

Figures 22 and 23 show the mean ammonia nitrogen and the sum of nitrite and nitrate nitrogen concentrations for the experiments conducted at 5°C. Ammonia nitrogen removal or conversion was insignificant (Figure 22). As mentioned above, some nitrification occurred in unit B. The increase in ammonia nitrogen concentration as the wastewater passed through the second and third stages in unit A was probably caused by lysis of cells. The plots in Figure 23 show that a small amount of nitrification occurred in units A, B, and C. Unit D operated at a liquid temperature of about 0.5°C less than the other three units, and this difference may have been critical for nitrification at this temperature. In addition, unit D was receiving the highest organic loading rate, and this may have inhibited the growth of nitrifiers. Since there was apparently no ammonia nitrogen removal although there was slight nitrification, it is possible that deamination of organic nitrogen followed cell lysis. Differences in the accuracy of the analytical methods were another possibility to explain the apparent nitrification. The straight parallel lines for nitrite-nitrate concentrations from the second through the fourth stages also indicate that the nitrification reaction followed zero order kinetics (Figure 23). The mean production rates for the sum of nitrite and nitrate nitrogen are presented in Table 26.

The mean for all of the values presented in Table 26 is 0.0235 $gN/m^2/d$, and without the extreme high

Table 23. Summary of steady-state mean nitrogen compounds concentrations in the mixed liquor in each stage of the RBC units operating at 5°C.

Unit	Parameter	First Stage		Second Stage		Third Stage		Fourth Stage	
		Mean	S.D.[b]	Mean	S.D.	Mean	S.D.	Mean	S.D.
B	F. Kjeldahl-N mg/l[a]	25.5	13.2	18.1	3.9	23.3	9.1	18.31	1.99
	NH_4-N mg/l	10.51	1.28	9.61	0.1	9.11	1.81	9.22	2.05
	NO_2-N mg/l	0.197	0.069	0.209	0.074	0.207	0.076	0.210	0.060
	NO_3-N mg/l	1.16	0.40	1.28	0.48	1.53	0.21	1.65	0.61
A	F. Kjeldahl-N mg/l	25.0	3.8	26.6	6.5	27.7	1.9	22.5	2.3
	NH_4-N mg/l	13.49	2.98	15.22	1.34	17.06	3.5	15.09	1.28
	NO_2-N mg/l	0.277	0.127	0.363	0.173	0.372	0.168	0.398	0.146
	NO_3-N mg/l	0.74	0.18	0.77	0.21	0.87	0.21	0.95	0.29
C	F. Kjeldahl-N mg/l	36.6	17.7	28.8	4.0	28.4	4.2	34.9	16.0
	NH_4-N mg/l	16.36	1.34	16.46	1.47	16.30	1.30	17.57	3.03
	NO_2-N mg/l	0.209	0.033	0.232	0.081	0.213	0.066	0.268	0.100
	NO_3-N mg/l	0.63	0.24	0.70	0.30	0.87	0.21	0.83	0.23
D	F. Kjeldahl-N mg/l	36.4	9.7	35.8	5.7	30.8	6.5	33.7	6.5
	NH_4-N mg/l	19.83	4.78	20.03	5.36	19.93	5.42	19.58	5.33
	NO_2-N mg/l	0.197	0.042	0.222	0.055	0.252	0.077	0.278	0.089
	NO_3-N mg/l	0.96	0.24	0.91	0.25	0.91	0.24	0.90	0.20

[a]Filtered Kjeldahl Nitrogen.

[b]Standard Deviation. The reported means were calculated from three separate analyses.

Table 24. Summary of steady-state mean nitrogen compounds concentrations in the mixed liquor in each stage of the RBC units operating at 15°C.

Unit	Parameter	First Stage			Second Stage			Third Stage			Fourth Stage		
		n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
A	TKN-mg/l	2	4.5	2.1	2	4.5	3.5	2	3.5	3.5	2	4.5	4.9
	Ammonia-N mg/l	2	1.05	1.05	2	0.59	0.4	2	0.44	0.45	2	0.40	0.39
	Nitrite-N mg/l	3	0.558	0.147	3	0.107	0.070	3	0.048	0.027	3	0.030	0.010
	Nitrate-N mg/l	3	7.94	1.00	3	8.62	1.94	3	9.17	1.56	3	9.57	1.64
B	TKN-mg/l	1	16.0	-	1	9.0	-	1	5.0	-	1	4.0	-
	Ammonia-N mg/l	2	14.58	6.47	2	4.00	4.85	2	2.03	2.69	1	0.36	-
	Nitrite-N mg/l	2	1.213	0.407	2	0.700	0.318	2	0.183	0.095	2	0.084	0.058
	Nitrate-N mg/l	2	2.04	1.82	2	11.30	1.03	2	12.32	1.15	2	12.92	1.48
C	TKN-mg/l	1	26.0	-	1	17.0	-	1	8.0	-	1	4.0	-
	Ammonia-N mg/l	2	18.15	0.46	1	9.64	-	2	1.39	0.38	2	0.43	0.33
	Nitrite-N mg/l	2	1.088	0.407	2	3.425	3.359	2	1.213	0.760	2	0.338	0.124
	Nitrate-N mg/l	2	0.06	0.01	2	10.07	0.17	2	15.79	2.88	2	18.54	2.78
D	TKN-mg/l	2	45.0	11.3	2	34.5	6.36	2	23.0	15.6	2	9.0	2.8
	Ammonia-N mg/l	2	29.33	2.59	2	23.65	0.74	2	13.34	4.94	2	3.89	2.94
	Nitrite-N mg/l	3	0.139	0.080	3	2.825	0.849	3	1.708	0.503	3	1.150	0.261
	Nitrate-N mg/l	3	0.27	0.046	2	2.79	2.07	3	10.29	3.14	3	20.93	3.05

Table 25. Summary of steady-state mean nitrogen compounds concentrations in the mixed liquor in each stage of the RBC units operating at 20°C (April 2 to April 12).

Unit	Parameter	First Stage			Second Stage			Third Stage			Fourth Stage		
		n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
A	F. Kjeldahl-N mg/l	2	12.15	1.41	3	1.16	0.65	3	0.70	0.64	3	0.50	0.27
	Ammonia-N mg/l	3	2.82	0.33	3	0.30	0.110	3	0.10	0.02	3	0.08	0.04
	Nitrite-N mg/l	3	1.215	1.472	3	0.770	0.328	3	0.333	0.090	3	0.195	0.082
	Nitrate-N mg/l	3	5.22	1.27	3	7.56	1.25	3	8.02	0.99	3	8.55	0.72
B	F. Kjeldahl-N mg/l	3	9.783	6.275	3	2.14	1.58	3	1.12	0.67	3	0.71	0.53
	Ammonia-N mg/l	3	5.92	0.99	3	1.17	0.97	3	0.44	0.36	3	0.21	0.16
	Nitrite-N mg/l	3	1.233	0.567	3	0.768	0.514	3	0.565	0.442	3	0.408	0.302
	Nitrate-N mg/l	3	3.35	0.85	3	8.08	1.64	3	8.90	0.36	3	9.32	0.45
C	F. Kjeldahl-N mg/l	3	20.05	5.334	3	7.57	4.57	3	2.53	0.89	3	1.43	0.36
	Ammonia-N mg/l	3	14.10	3.03	3	1.66	1.00	3	0.31	0.14	3	0.08	0.03
	Nitrite-N mg/l	3	0.617	0.302	3	1.307	0.352	3	0.587	0.304	3	0.320	0.214
	Nitrate-N mg/l	3	1.55	0.61	3	10.74	1.63	3	13.50	1.69	3	13.83	1.64
D	F. Kjeldahl-N mg/l	3	22.28	9.24	3	7.17	8.53	3	3.39	2.45	3	2.96	1.75
	Ammonia-N mg/l	3	14.85	1.36	3	3.23	3.17	3	1.34	1.58	3	1.50	1.50
	Nitrite-N mg/l	3	0.567	0.489	3	0.925	0.522	3	0.775	0.538	3	0.648	0.538
	Nitrate-N mg/l	3	2.02	1.01	3	10.99	3.08	3	13.72	2.75	3	13.64	3.48

Table 26. Nitrification rates at 5°C ($gN/m^2/d$).

Unit	Stage 2	Stage 3	Stage 4
A	0.0222	0.0226	0.0195
B	0.0258	0.0482	0.0226
C	0.0174	0.0297	0.0036

and low values, the mean is 0.0228 $gN/m^2/d$ with a standard deviation of 0.0040 $gN/m^2/d$. Figure 24 shows the mean steady-state ammonia nitrogen concentrations in each stage of the experimental RBC units operating at 15°C. Figures 25 and 26 present the mean steady-state ammonia nitrogen concentrations when the units were operating at 20°C for the first (March 20 - March 27) and the second sampling periods (April 2 - April 12), respectively. Generally in the first stages there was limited ammonia nitrogen removal except in unit A, which was receiving the lowest organic loading rate. Significant ammonia nitrogen removal occurred in the second stages, and proceeded in the following stages in the units receiving high organic loading rates. The declining ammonia nitrogen removal rates were observed in the stages containing low concentrations of ammonia nitrogen and indicate substrate limiting conditions. In the region where substrate was not limiting, the decline in ammonia nitrogen removal followed a straight line, and the lines for the different units are generally parallel. At a temperature of 20°C, the slopes of these lines are greater than the slopes of the lines at 15°C emphasizing the effect of temperature on ammonia nitrogen removal rates beyond 15°C.

Based upon the above observations, it appears that ammonia nitrogen removal could be described by Michaelis-Menten enzyme kinetics. An analysis of the data using Michaelis-Menten enzyme kinetics is presented in another section of this report. Table 27 contains a summary of the ammonia nitrogen removal rates when substrate was not limiting. These rates can be considered the maximum removal rates. The mean maximum removal rate at 15°C was 1.956 $gN/m^2/d$ and 2.694 $gN/m^2/d$ at 20°C.

Table 28 presents a summary of the overall ammonia nitrogen removal efficiencies at 15°C and 20°C. Only the second sampling period data were considered from the 20°C experiments because an adequate data base was not available during the first sampling period. The results in Table 28 show that 98 to 99 percent ammonia nitrogen removal was obtained at organic loading rates up to 10 to 12.5 $gCOD/m^2/d$. The removal efficiency decreased by approximately 10 percent at organic loading rates of 14 $gCOD/m^2/d$.

Table 27. Summary of maximum ammonia removal rates for each stage of the RBC units operating at temperatures of 15°C and 20°C ($gN/m^2/d$).

Unit	Temperature 15°C			Temperature 20°C	
	Stage 2	Stage 3	Stage 4	Stage 2	Stage 3
B	2.195	-	-	2.651	-
C	1.745	1.692	-	2.739	-
D	-	2.164	1.984	3.301	2.713
Second period for 20°C					
C				2.441	-
D				2.318	-

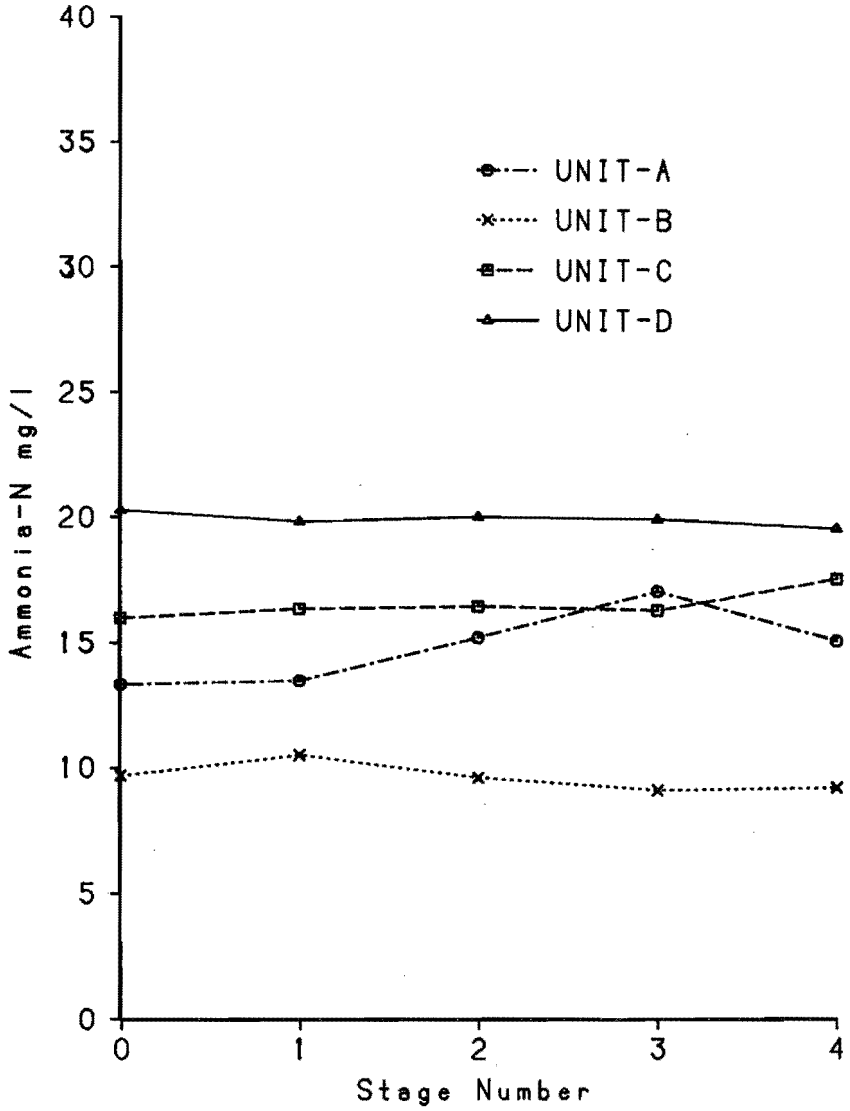


Figure 22. Mean steady-state mixed liquor ammonia nitrogen concentrations in the four stages of the RBC units operating at 5°C.

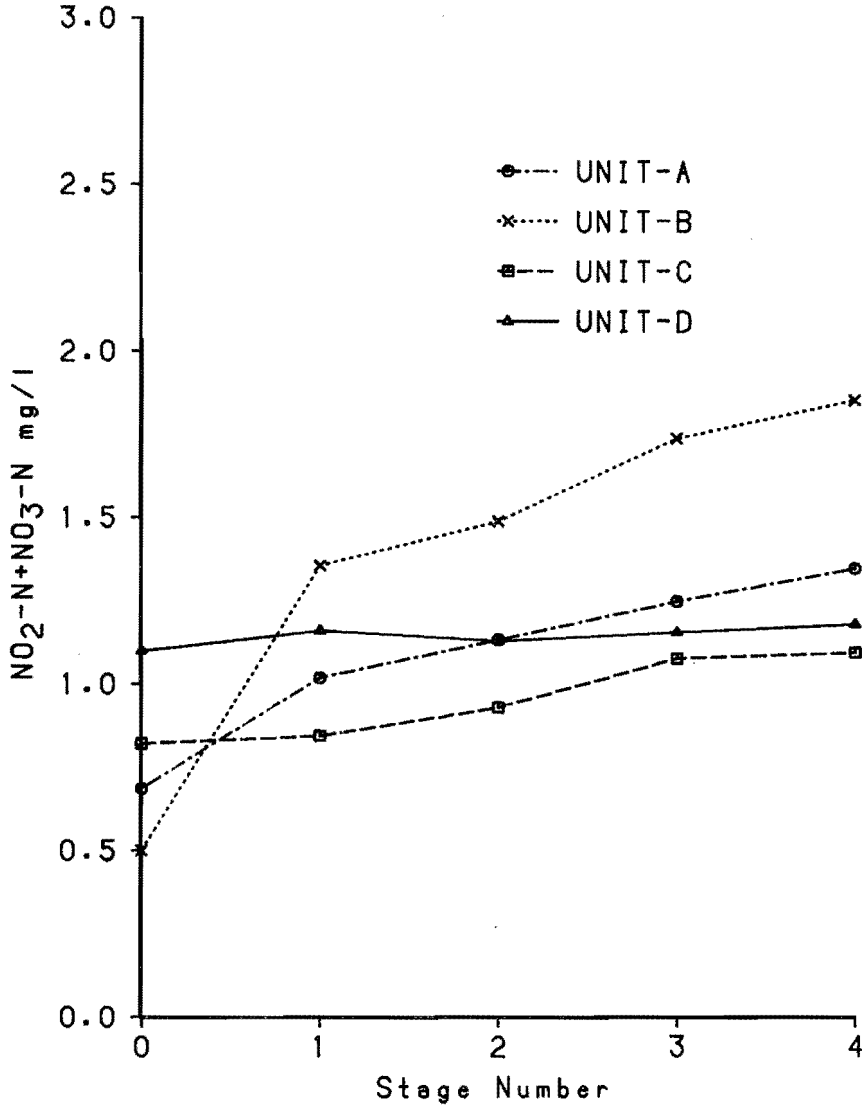


Figure 23. Mean steady-state mixed liquor nitrite and nitrate nitrogen concentrations in the four stages of the RBC units operating at 5°C.

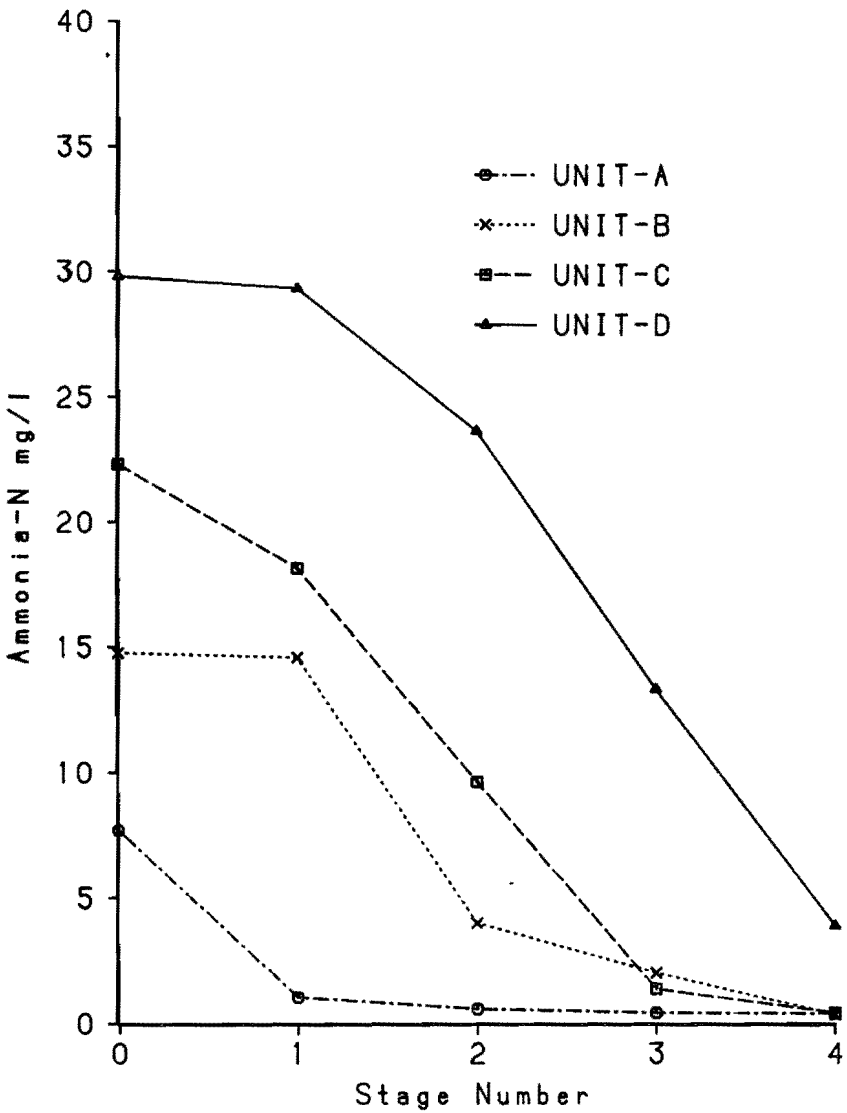


Figure 24. Mean steady-state ammonia nitrogen concentrations in the four stages of the RBC units operating at 15°C.

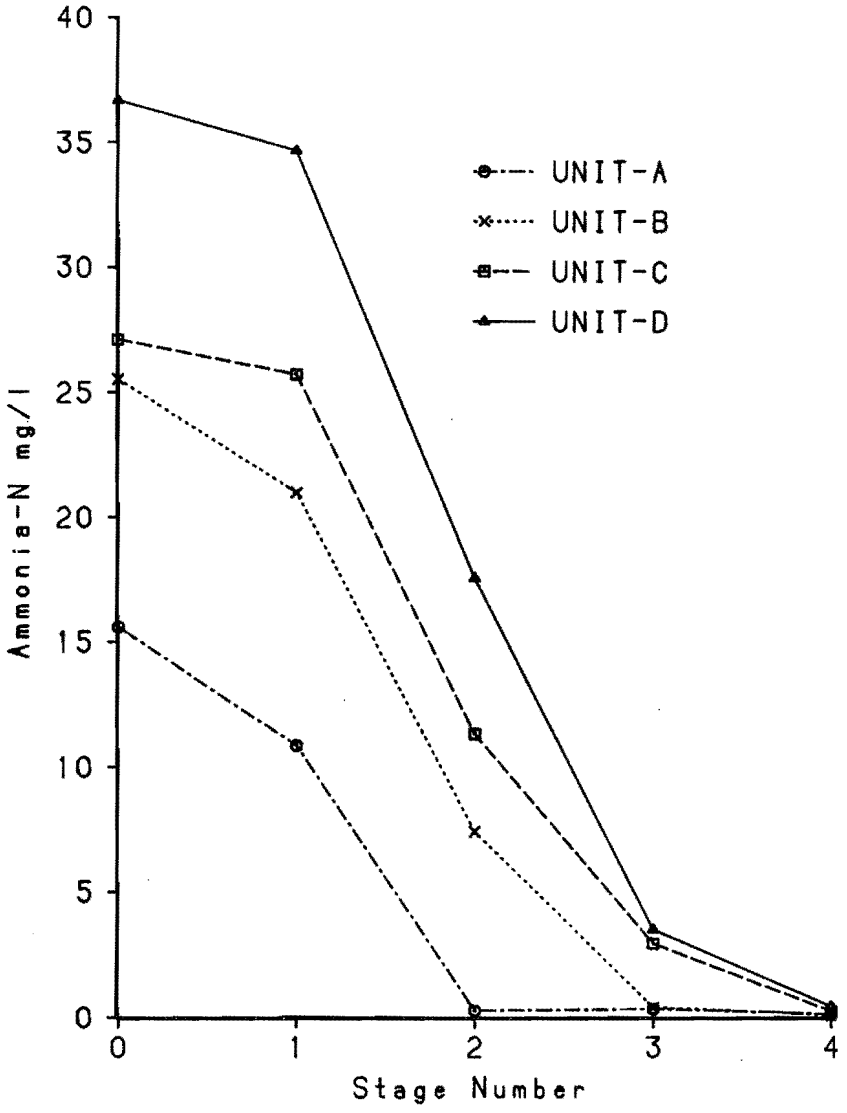


Figure 25. Mean steady-state ammonia nitrogen concentrations in the four stages of the RBC units operating at 20°C (first period).

The percentage removal of ammonia nitrogen was higher at 20°C (Table 28). Figures 27, 28, and 29 present the sum of nitrite and nitrate nitrogen concentrations in the stages of the RBC units at 15°C and at 20°C during the first and second sampling periods, respectively. The trends in these figures generally show the same pattern in nitrification rate as that observed for the ammonia nitrogen removal rate.

The nitrite nitrogen concentrations through the stages demonstrate specific characteristics that were probably associated with the environmental conditions in which the nitrifiers were growing. At both 15°C and 20°C, the nitrite nitrogen concentrations peaked between the influent and the third stage (Figures 30 and 31). In units A and B peaks of approximately 1 mg/l nitrite nitrogen were observed in the first stage. In units C and D the peaks occurred in the second stage where significant nitrification was observed. Comparing the amount of nitrification to the amount of ammonia nitrogen removed, it appears that there was more nitrification in unit A than in the other units. The reason could be a combination of processes, including nitrification, ammonia nitrogen assimilation and stripping, and ammonia nitrogen release into the bulk solution through cell lysing following deamination of organic nitrogen. Units receiving low organic loading rates experienced more cell lysing as has been shown earlier when examining the carbonaceous substrate characteristics through the stages. However, in both cases, the differences in absolute values were not significant. The contribution of nitrification to ammonia nitrogen removal in the RBC units is summarized in Table 29. Denitrification can contribute to losses of nitrogen, but it does not appear to contribute to ammonia nitrogen removal in these RBC units operating under aerobic conditions. Table 29 shows that more than 85 percent (mean value of 90 percent) of ammonia nitrogen removal is attributable to the nitrification process in units B, C, and D. The relatively high deviation obtained in unit A at 15°C was probably due to the very low organic loading rates applied to unit A at 15°C as shown in Table 28. Consequently, ammonia nitrogen release into the solution was stimulated by deamination of organic nitrogen following cell lysing.

Table 28. Summary of organic and ammonia nitrogen loading rates and the ammonia nitrogen removal efficiency.

Temperature °C	Unit	Organic Load, g COD/m^2/d	Ammonia Load g N/m^2/d	Removal Efficiency Percent
15	A	3.984	0.387	94.8
	B	7.496	0.766	97.6
	C	9.875	1.143	98.1
	D	13.916	1.563	86.9
20	A	6.915	0.362	99.0
	B	9.734	0.520	98.0
	C	12.513	0.663	99.4
	D	13.971	0.786	90.5

Table 29. Summary of the percentage of ammonia nitrogen removal attributable to nitrification (%) in the RBC units.

Temperature °C	Unit A	Unit B	Unit C	Unit D
15	131	90	86	85
20	112	87	101	95

Attached biomass characteristics

At the end of each phase of this study, the total amount of biomass attached to the discs in each stage was measured. In the first phase, the biomass on each disc in all four stages was measured separately, and resulted in uniform amounts per disc when analyzed by stage. In all three phases, there was a successive decrease in the quantity of biomass attached to the discs from the first to the fourth stages. At lower organic loading rates and higher temperatures, there was a sharp decline in the quantity of attached biomass following the first or second stage. At lower organic loading rates and higher temperatures, less substrate and less unstabilized sloughed biomass were available to establish attached growth in these later stages. Table B-31 (Appendix B) contains a summary of the measurements of the biomass attached to the discs and the volatile content of the solids. The following discussion considers the biomass as volatile solids.

Figures 32, 33, and 34 show the variation in the quantity of attached biomass in the four stages of the RBC units operating at 5°C, 15°C, and 20°C, respectively. The attached biomass increased in the higher loaded units until a maximum was reached. This observation supports the contention that the mass of attached growth is dependent upon the organic loading rate and can be defined by a saturation function. As the temperature increased, the maximum mass of attached biomass increased linearly ($R^2 > 0.996$) with the equation of the line having a slope of 1.05 grams VS/m^2 per 1°C in a temperature range of 6°C to 21°C for the first stages. This linear relationship indicates that the maximum growth per unit disc area was not governed by the physical capability of the discs to carry the biomass, but by the biological activity. Generally there was a gradual decline in the quantity of biomass from stage to stage. The rate of decline increased as the temperature increased indicating less available substrate and less unstabilized sloughed biomass to serve as substrate to produce more attached growth. Attached growth develops in the second, third, and fourth stages although further reduction in the soluble substrate does not occur. While in the first stage, most of the available substrate was consumed; in the later stages the sloughed biomass from the preceding stages was reattached and stabilized. While stabilization increased (less stored substrate in the cell material) as the wastewater flowed through the stages, the reattachment decreased resulting in less attached biomass. As the temperature increased, this process was stimulated and there was a decline in attached biomass in the earlier stages.

The maximum attached biomass per unit area in terms of VSS in the first stages were 24.9 g/m^2, 36.2 g/m^2, and 40.5 g/m^2 at 5°C, 15°C, and 20°C, respectively. In terms of total solids per area, the corresponding values were 37.9 g/m^2, 55.4 g/m^2,

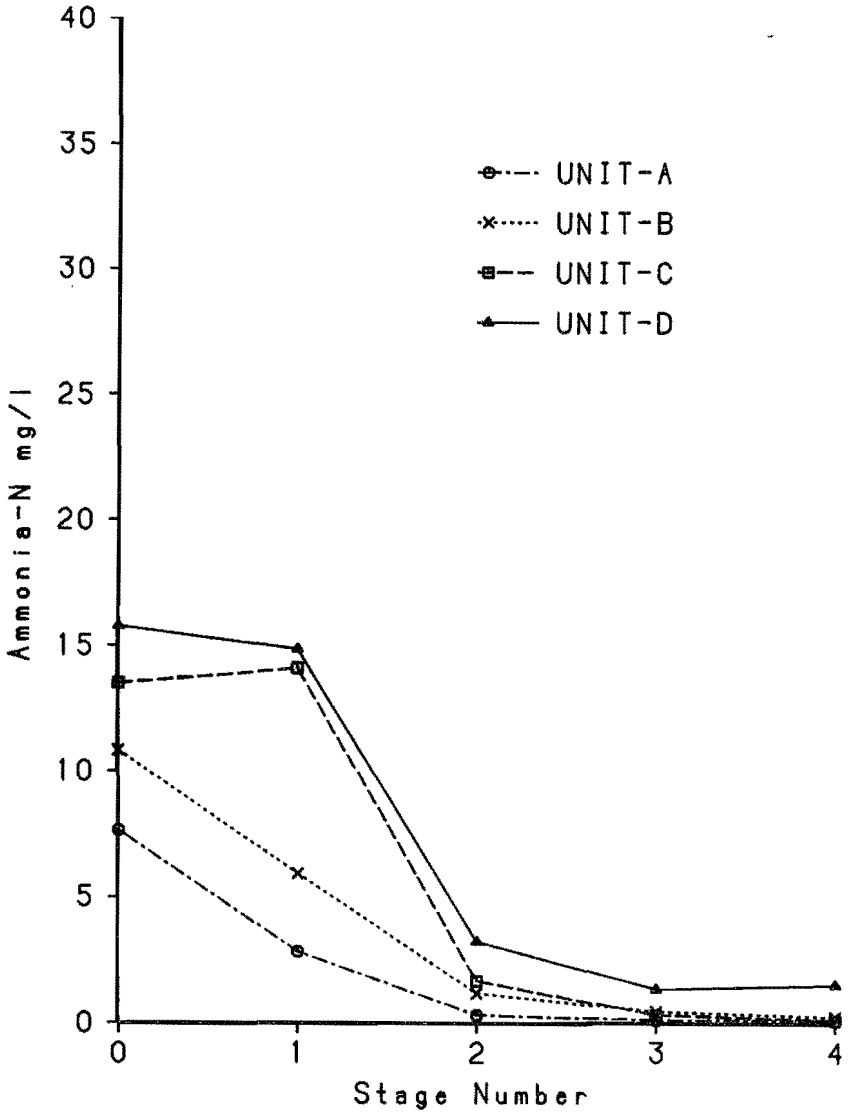


Figure 26. Mean steady-state ammonia nitrogen concentrations in the four stages of the RBC units operating at 20°C (second period).

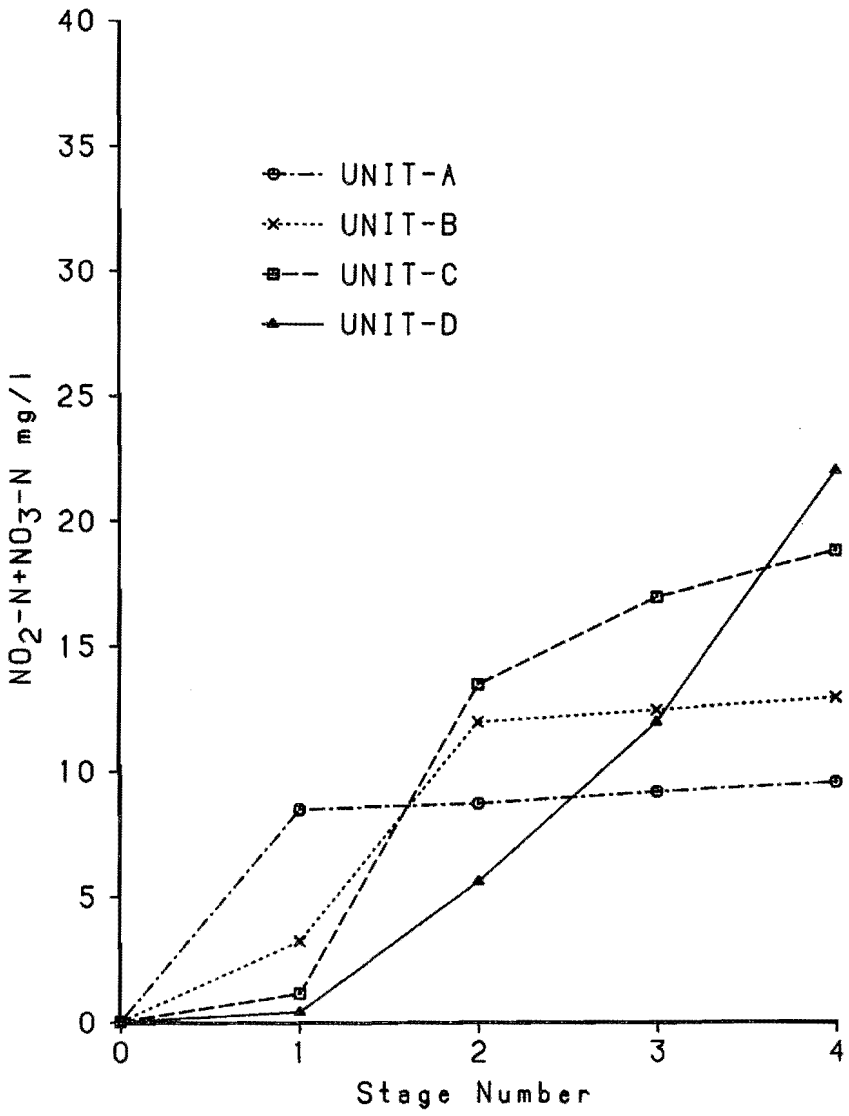


Figure 27. Mean steady-state nitrite and nitrate nitrogen concentrations in the four stages of the RBC units operating at 15°C.

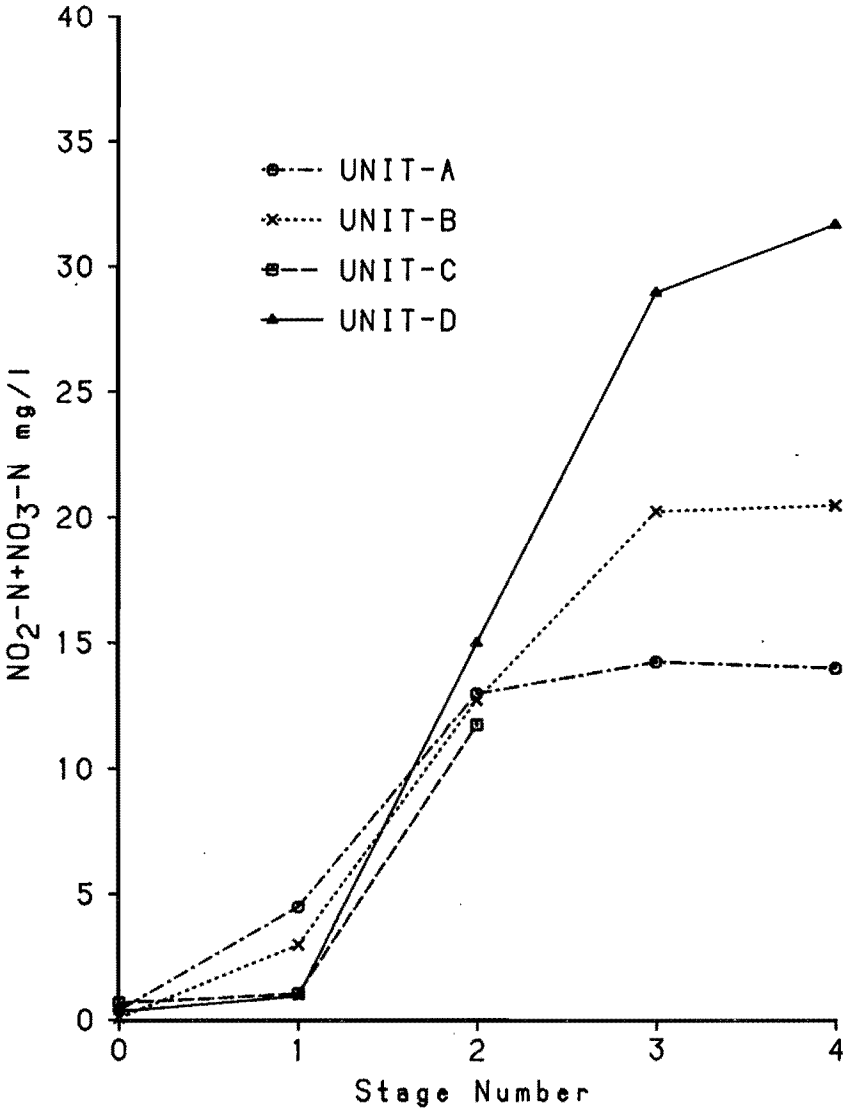


Figure 28. Mean steady-state nitrite and nitrate nitrogen concentrations in the four stages of the RBC units during the first sampling period with the units operating at 20°C.

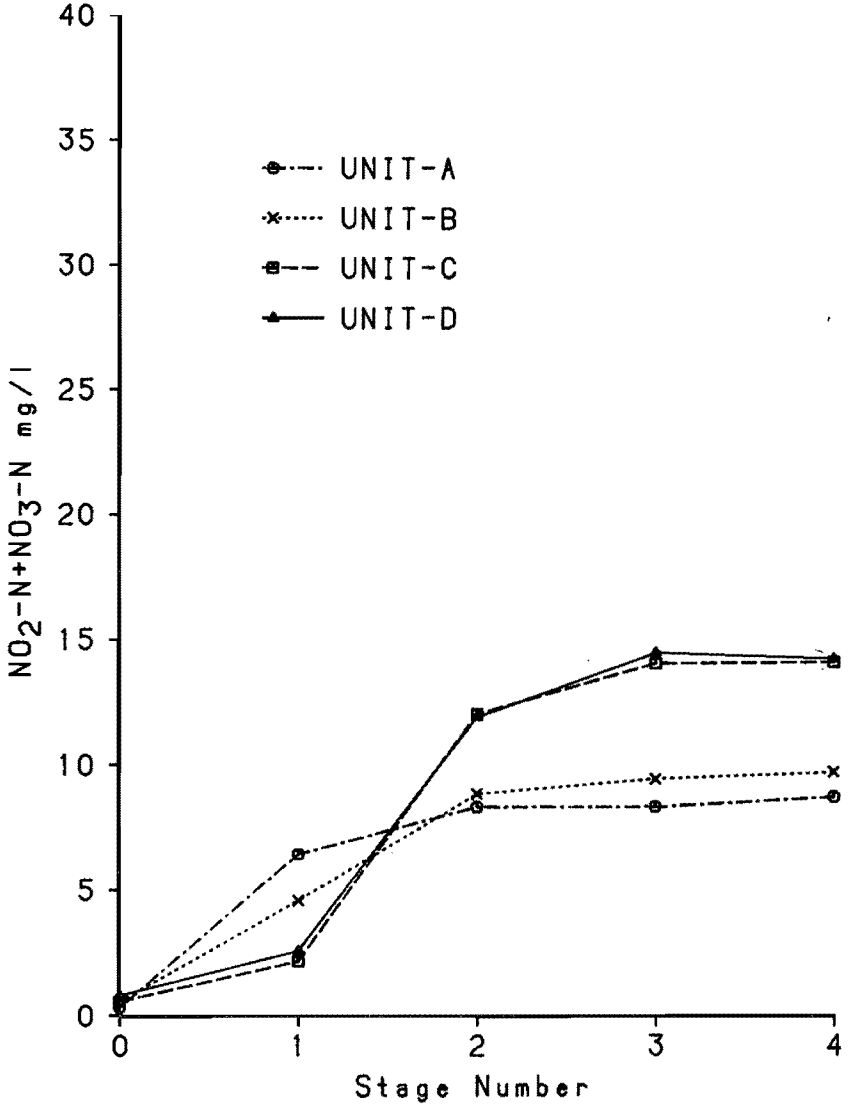


Figure 29. Mean steady-state nitrite and nitrate nitrogen concentrations in the four stages of the RBC units during the second sampling period with the units operating at 20°C.

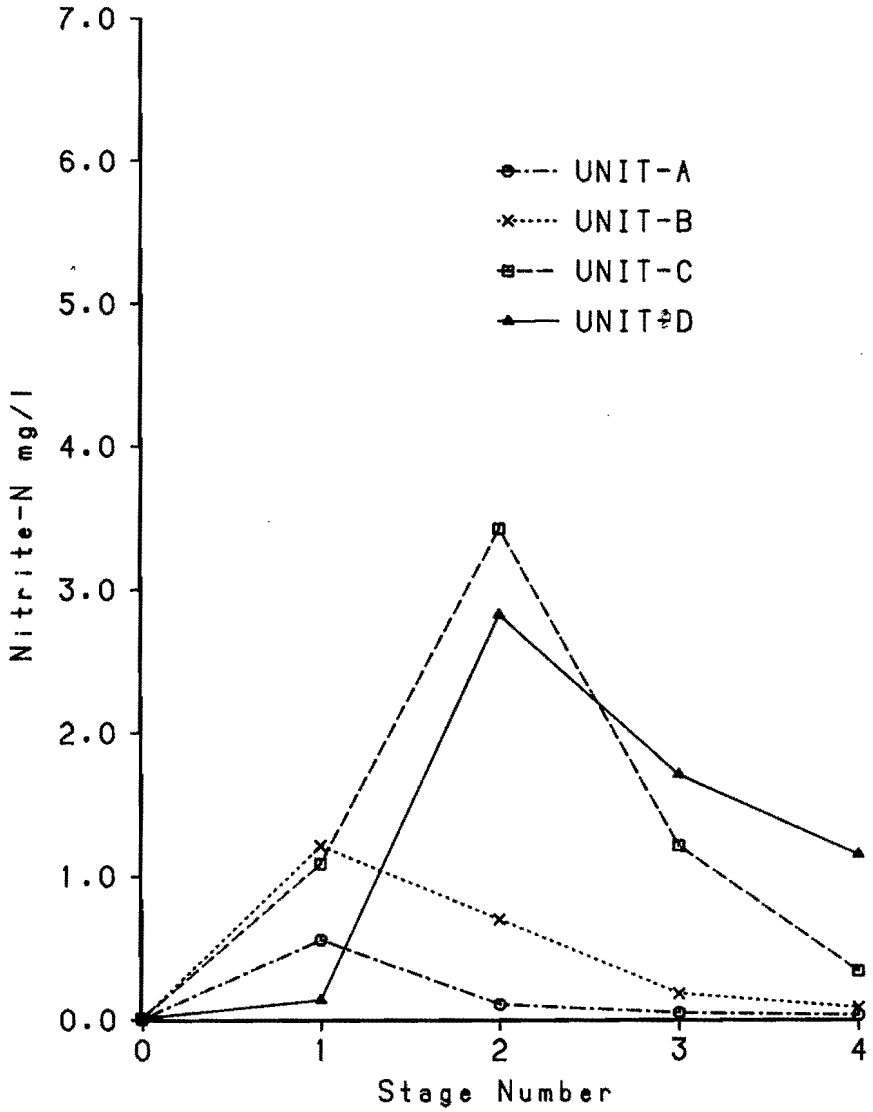


Figure 30. Mean nitrite nitrogen concentrations in the four stages of the RBC units operating at 15°C.

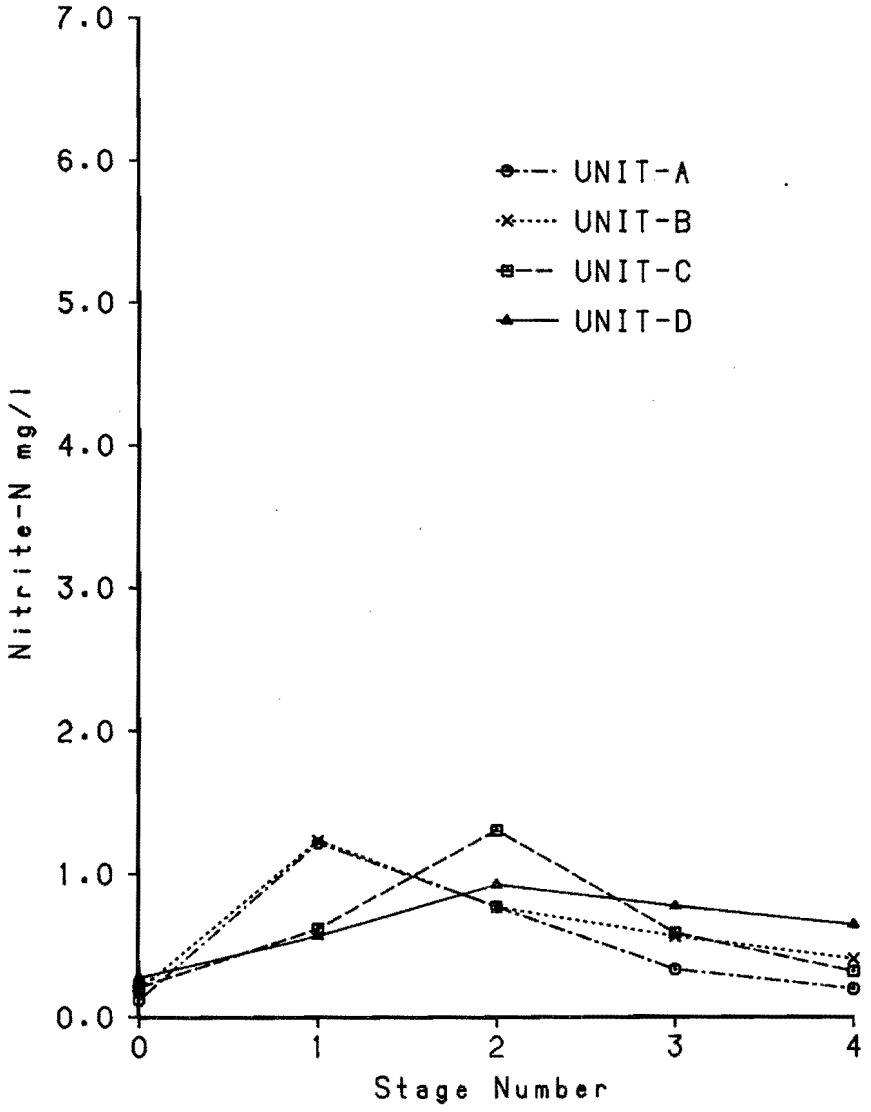


Figure 31. Mean nitrite nitrogen concentrations in the four stages of the RBC units operating at 20°C.

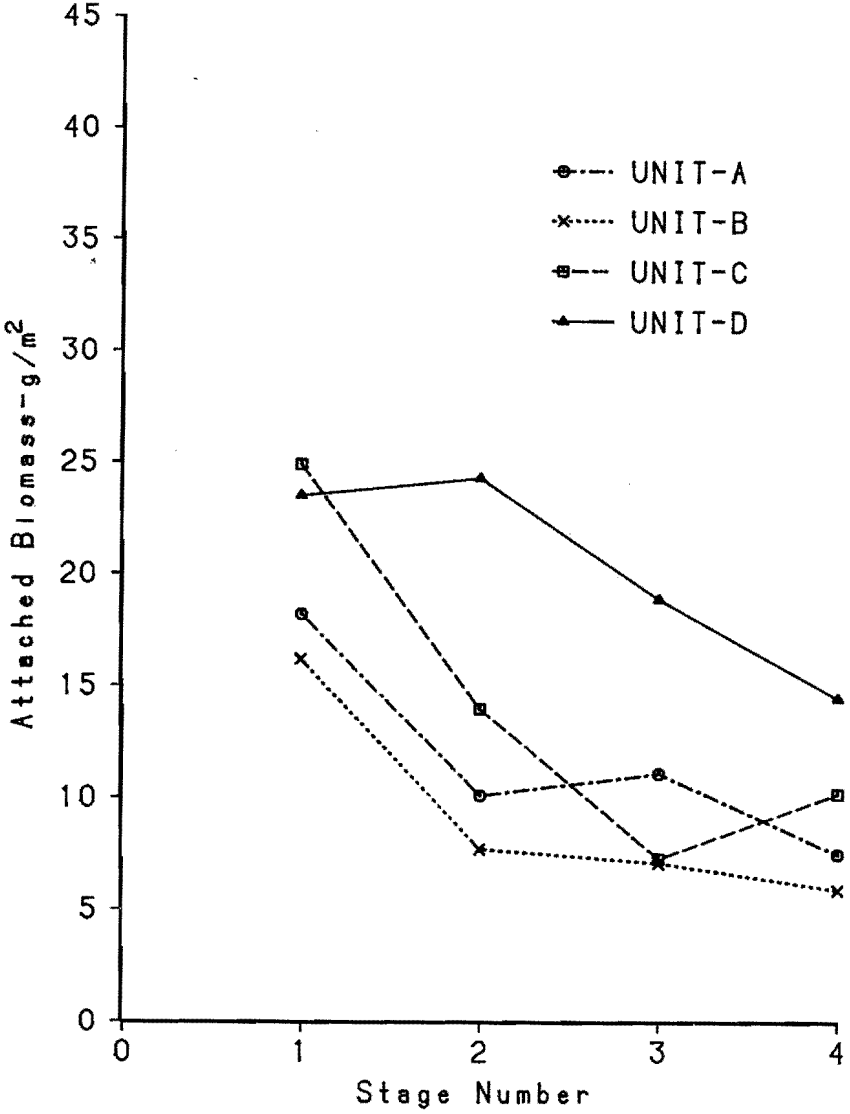


Figure 32. Attached biomass in the four stages of the RBC units operating at 5°C

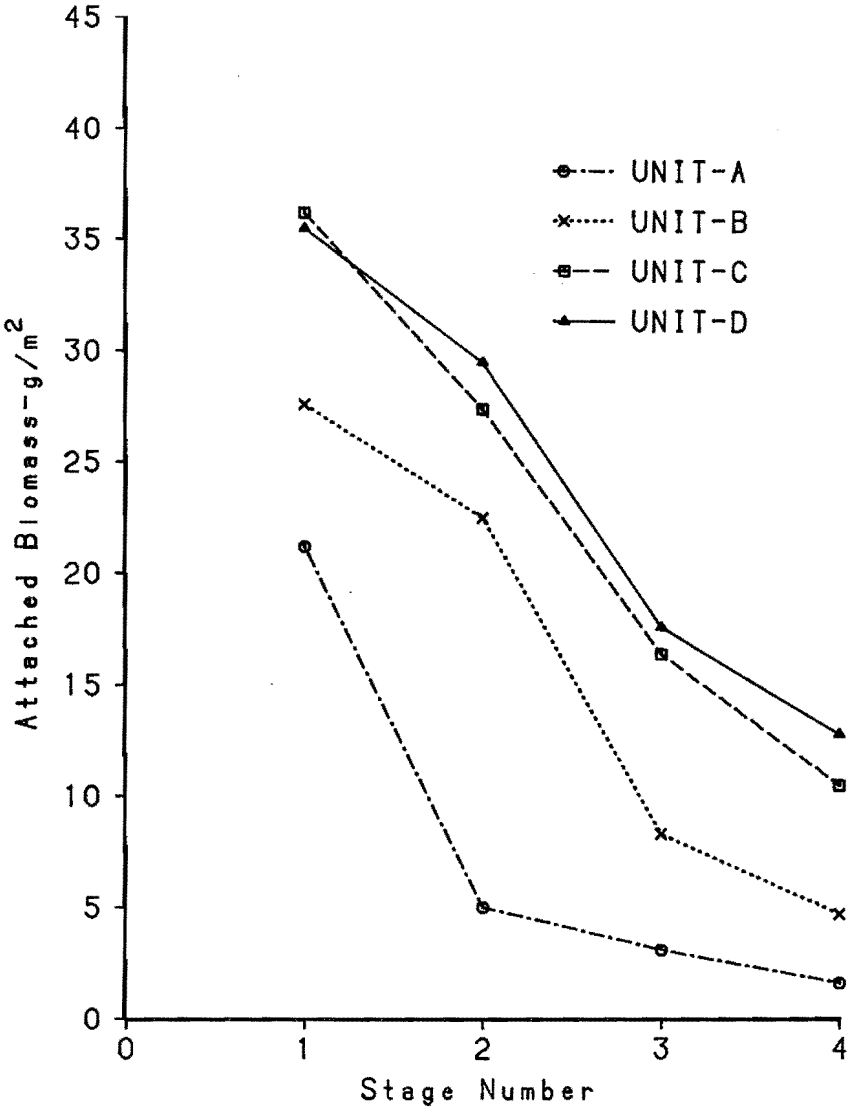


Figure 33. Attached biomass in the four stages of the RBC units operating at 15°C.

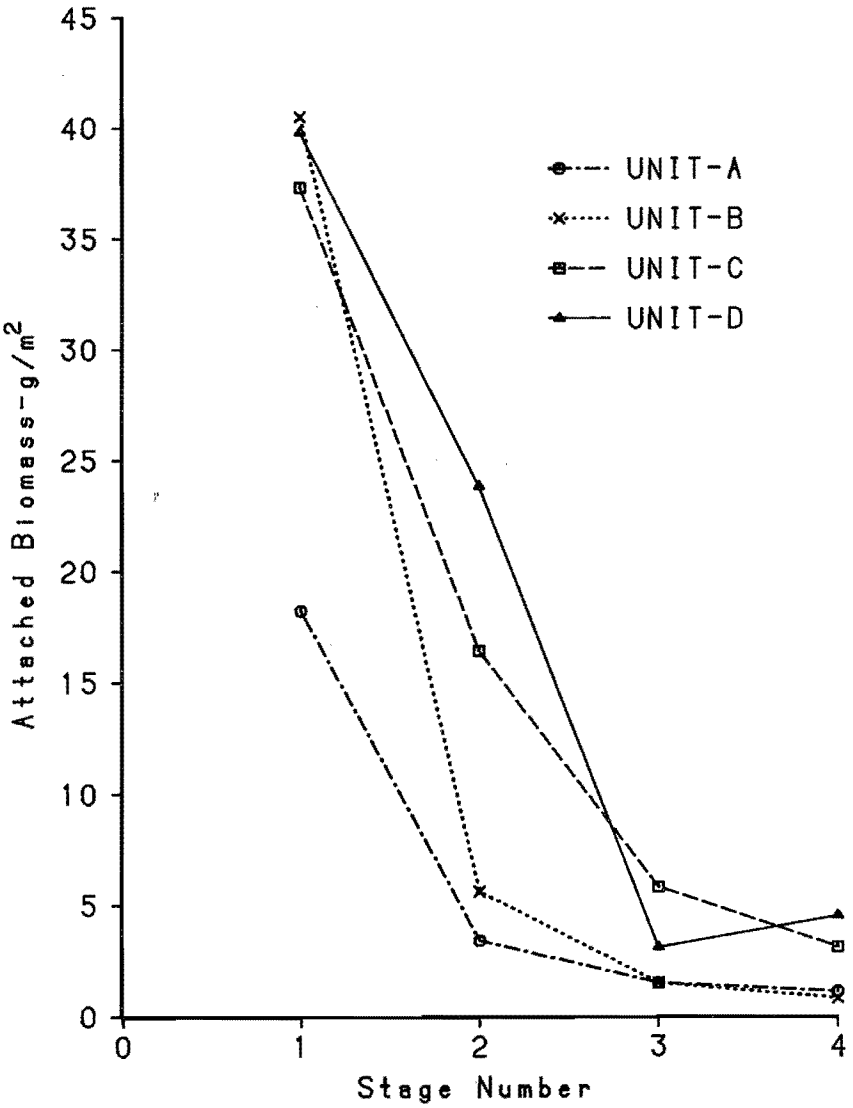


Figure 34. Attached biomass in the four stages of the RBC units operating at 20°C.

and 60.7 g/m^2. Maximum values reported in the literature are 45 g/m^2 (Pretorius 1971), 65 g/m^2 (Ouyang 1980), and 51 g/m^2 (Clark et al. 1978).

Figure 35 shows the relationships between the first stage biomass and the organic loading rate at the three temperatures. The relationships show a saturation type function and the effect of temperature.

A summary of the VS fraction of the attached biomass for all the three phases is shown in Table B-32 (Appendix B). Table 30 presents the mean values for these phases. As shown in Table 30, there is a trend toward a decline in the VS fraction from the first stages to the last stages. This indicates that the biomass is more highly stabilized as the wastewater progresses through the stages. The chemical analyses performed on samples of biomass at 5°C and 15°C are presented in Table 31. At 5°C the organic carbon and nitrogen content of the attached biomass was less than the values obtained at 15°C. This may be due, among other reasons, to less stored substrate in the cells at 5°C than at 15°C because of less biological activity. In the first stages, the percentage of the attached biomass composed of TOC were 32.2 and 41.1 percent for 5°C and 15°C, respectively. The total nitrogen in the attached biomass of the first stages was 6.40 and 7.65 percent at temperatures of 5°C and 15°C. The ratio of carbon to nitrogen was 5.4 to 5.2 at these temperatures. Ouyang (1980) reported a C to N ratio of 7.0, and Kornegay and Andrews (1968b) reported a ratio of 4.3 using glucose as the substrate. At 15°C in this study, the weight ratio of C:N:P was 28:5.2:1. A formulation for bacteria has been proposed by McCarty (1970) to be $C_{60}H_{87}O_{28}N_{12}P$, which results in a weight ratio of 23.2:5.4:1. On a volatile fraction basis the organic carbon contents were 48 percent and 59 percent for temperatures of 5°C and 15°C, respectively.

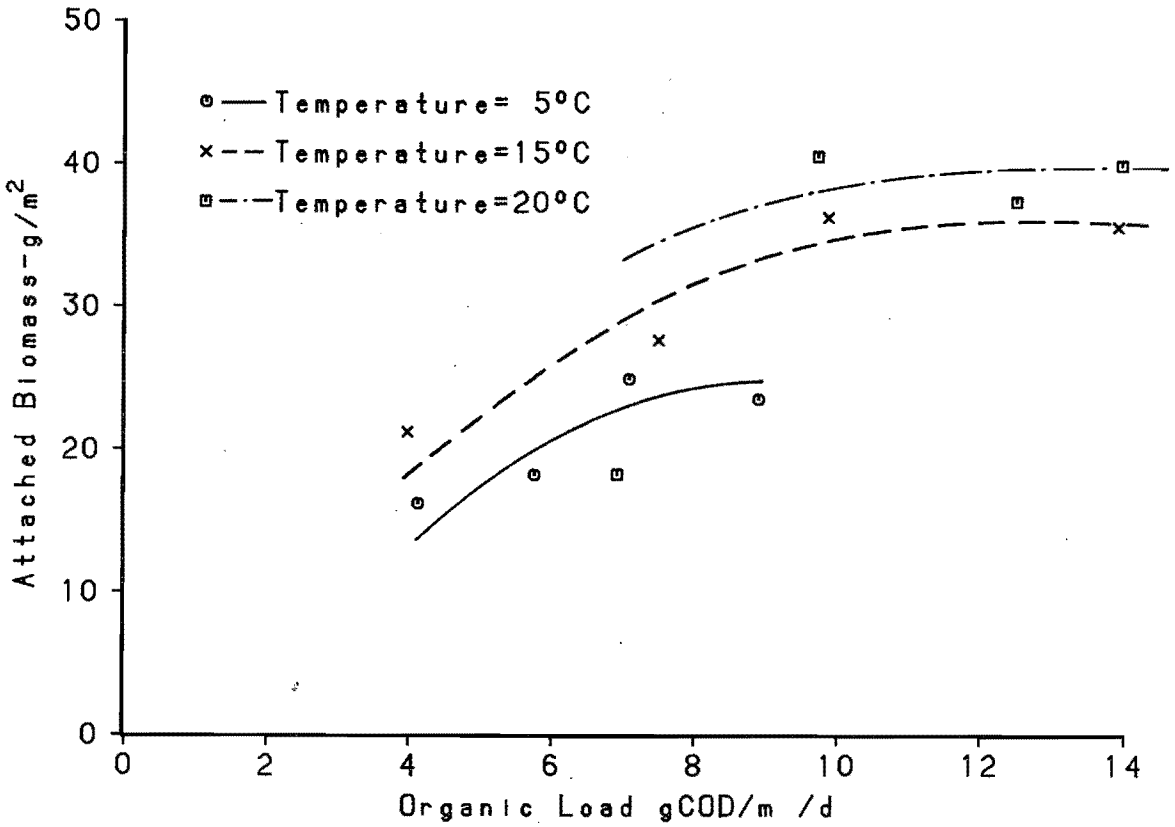


Figure 35. The relationship between temperature, organic load, and the mass of attached biomass per unit of surface area in the first stages of the RBC units.

Table 30. Summary of the mean values of the VS fraction in the attached biomass (%).

Stage Number	5°C		15°C		20°C	
	Mean	Range	Mean	Range	Mean	Range
1	67.11	64.29-71.11	69.52	60.25-82.05	66.95	66.0-68.1
2	64.62	63.48-65.41	62.50	54.71-66.53	67.40	64.0-73.7
3	64.42	62.81-65.98	67.64	58.31-79.28	60.38	48.9-64.8
4	62.31	58.79-64.33	58.16	53.83-64.43	62.89	58.8-67.7

Table 31. Summary of chemical composition of the attached biomass (%) (dry basis).

Unit State	5°C		15°C		Total P
	TOC	Total-N	TOC	Total-N	
A-1	33.2	6.49	39.6	7.16	1.359
B-1	29.2	6.50	42.3	7.33	1.429
C-1	34.5	6.26	42.8	7.98	1.545
D-1	31.7	6.33	39.8	8.11	1.545
C-2	35.4	6.53	39.5	7.58	1.405
D-2	33.3	6.36	36.3	7.26	1.533
C-3	-	-	34.9	6.75	1.452
D-3	30.8	6.12	38.2	6.98	1.521
D-4	39.2	5.91	36.9	6.52	1.719

KINETIC MODEL DEVELOPMENT

General Considerations

Kinetic models based on rational considerations were developed, and the kinetic coefficients were estimated independently for each temperature. Only reaction expressions for biomass growth or substrate removal containing a maximum of two kinetic constants at a specific temperature were considered. The data collected from the four experimental RBC units were adequate to determine the kinetic constants independently with two degrees of freedom for each temperature. Consequently, the temperature dependence of the kinetic constants could also be examined.

The determinations of the kinetic constants were based upon the mean steady-state values of the parameters measured for each unit. The concentrations of the pollutants in the effluent were independent of the fluctuations in the influent concentrations; therefore, the mean concentrations from each unit were utilized in the calculations. As an example of this independence, the deviations around the mean of the effluent substrate fluctuations and the influent substrate fluctuations of the first stage of unit D are shown in Figure 36 for filtered COD. The same behavior was observed for the other units and other pollutants.

Because the first stage of the RBC units performed differently for carbonaceous substrate removal from the other stages, separate estimates of the kinetic constant were made for this stage. A distinguishing feature of the first stages was that they were receiving raw wastewater, while the influent to the other stages contained sloughed biomass from the preceding stages and unconsumed substrate. This difference in substrate affects the processes taking place in the RBC stages. In the first stages most of the carbonaceous substrate is consumed, while in the following stages nitrification proceeds, sloughed biomass is stabilized, and there is further consumption of the substrate in the mixed liquor. Differences in the processes also are apparent in the appearance of the attached biomass. In the first stages, the biomass is spongy and relatively stable. The major processes that can be related to the biomass in the first stages are the carbonaceous substrate removal and endogenous respiration of the attached growth. In the following stages the attached biomass has a smooth appearance and is associated with several processes, i.e., stabilization of reattached biomass, nitrification, substrate consumption and endogenous respiration. Consequently, in the kinetic model development for carbonaceous substrate removal, it can be justified in the first stages to consider all of the

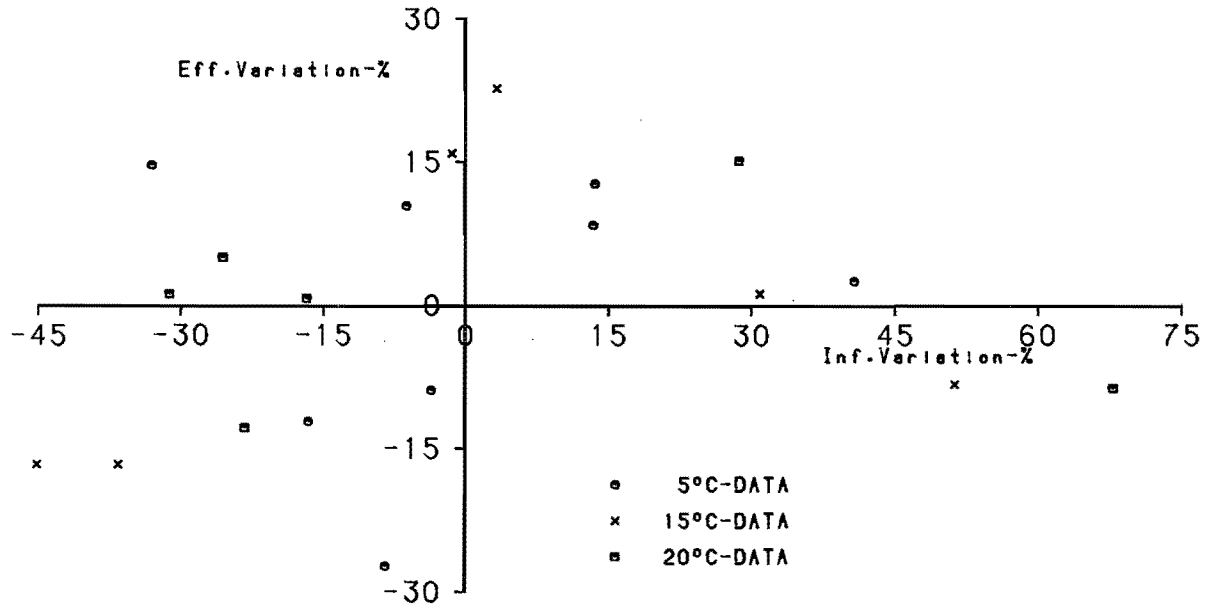


Figure 36. Variation of influent and effluent filtered COD in the RBC unit D.

attached biomass. In the second through the fourth stages, that cannot be done. Therefore, the carbonaceous substrate removal kinetics were developed separately for the first stages than for the other stages of the RBC process.

In the second through the fourth stages, an instability was observed in the mixed liquor characteristics for the suspended solids and the carbonaceous substrate, making it difficult to develop a mathematical model for each stage. To compensate for this instability, carbonaceous substrate removal kinetics and sloughed biomass stabilization kinetics for the three stages were considered as one reactor. The similarity of the processes in these stages justify the use of this approach.

In ammonia nitrogen removal kinetics model development, the data for all of the stages where nitrification was not inhibited are considered as independent steady-state data. As a result there are more degrees of freedom in the mathematical model development for ammonia nitrogen removal. The same rates of ammonia nitrogen removal were observed in all of the stages where nitrification was not restricted and substrate limitations did not occur. Therefore, a single kinetic model appears reasonable regardless of stage number. Kinetic models for the following processes are presented in the following sections of this chapter:

1. Biomass yield and decay rates.
2. Carbonaceous substrate removal in the first stages.
3. Carbonaceous substrate removal in the second through the fourth stages.
4. Sloughed biomass stabilization in the second through the fourth stages.
5. Ammonia nitrogen removal.

Biomass Yield and Decay Rates

General assumptions

Biomass yield and decay rates for carbonaceous substrate removal could be evaluated reliably only in the first stages of the RBC units. The stability of the attached biomass in the first stages lead to the assumption that the quantity of attached biomass was constant during the steady-state period. All attached biomass losses, except sloughing, were combined in the endogenous decay rate. Moreover, as shown for other systems such as the activated sludge process, the decay rate is related to the quantity of biomass measured as volatile solids without distinguishing between active and nonactive biomass (Schroeder 1977; Muck and Grady 1979).

The particulate material in the mixed liquor of the RBC stages was assumed to be the sloughed biomass from the attached growth. The remaining substrate in the stage effluent was the filtered substrate. The substrate in the influent available for growth was the total substrate which includes the filtered and the colloidal substrate.

For the experiments conducted at 15°C and 20°C with units receiving lower organic loading rates, the biomass yield associated with the 5 to 7 mg/l of ammonia nitrogen removed in the first stages was neglected. Applying a yield coefficient of 0.15 for nitrifiers (Ito and Matsuo 1980), the yield resulting from 5 to 7 mg/l of ammonia nitrogen removed would be approximately 1 mg/l VSS. This is less than 0.3 percent of the total attached biomass.

Substrate removal by the suspended biomass in the mixed liquor and the yield associated with it were also neglected. The maximum concentration of VSS (143 mg/l) in the first stages was obtained with unit D at 15°C. Using the following set of kinetic constants for suspended growth (Metcalf and Eddy, Inc. 1979):

$$k = 5/d;\ K_s = 40\ mg/l\ COD;\ Y = 0.4,$$

and substituting the mean steady-state data for unit D into Equation 31 results in an estimate of COD removal of 1.4 mg/l. The associated biomass yield is less than 1 mg/l

$$Q\Delta S = V_1 \frac{k\ X_1\ S_1}{K_s + S_1} \qquad (31)$$

where

Q = Flow rate = 288.6 l/d for unit D at 15°C

S = Substrate removed, mg/l

V_1 = Volume of the first stage = 6 liter (in the first phase)

X_1 = Mixed liquor VSS = 143 mg/l for unit D at 15°C

S_1 = First stage effluent filtered COD = 49.7 mg/l for unit D at 15°C

k = Maximum reaction rate, 1/d

K_s = Half saturation constant, mg/l COD

Y = Yield coefficient, mg VSS/mg COD removed

Kinetic model for biomass yield and decay

In rotating biological contactors (RBC), simultaneous reactions and mass transport processes occur both within the liquid and in the attached growth (semisolid phase). Figure 37 shows both processes schematically in a flow diagram for the first stage of RBC treatment.

A mass balance of the attached growth can be derived as follows:

Accumulation = (Influent biomass)-(sloughed biomass) + (biomass growth rate)-(biomass decay) (32)

The rates of biomass growth and substrate removal can be related with a yield coefficient defined as:

$$\text{Yield} = \frac{\text{biomass growth rate}}{\text{substrate removal rate}} \qquad (33)$$

Incorporating the yield coefficient into Equation 32 and using Figure 37 to define the other terms gives:

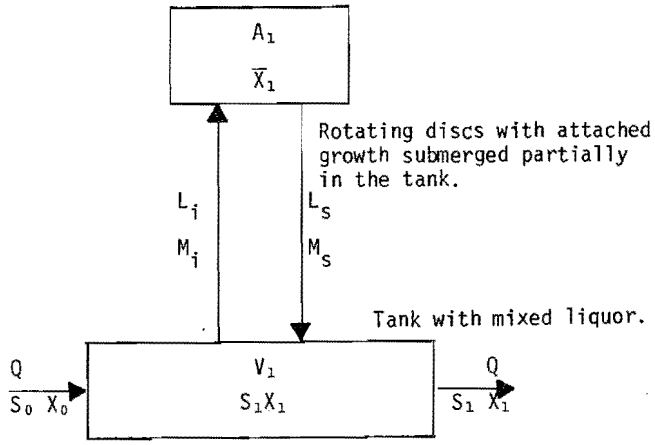
$$A_1 \frac{d\overline{X}_1}{dt} = M_i - M_s + Y\,(L_i - L_s) - k_d A_1 \overline{X}_1 \qquad (34)$$

where

t = time, day

Y = yield coefficient

k_d = decay rate, 1/d.



A_1 = Total available surface area for attached growth, m^2

$\overline{X}_1$ = Attached growth per unit area, grams/m^2

Q = Influent wastewater flow rate, m^3/d

V_1 = Mixed liquor volume in the tank, m^3

S_0 = Influent substrate concentration, mg/l

S_1 = Effluent substrate concentration, mg/l

X_0 = Influent microorganism concentration = 0

X_1 = Effluent microorganism (biomass) concentration, mg/l

L_i = Substrate adsorbed and absorbed from the mixed liquor by the rotating attached growth, grams/d

L_s = Substrate released from the rotating attached growth into the mixed liquor, grams/d

M_i = Mixed liquor biomass entrapped within the rotating discs, grams/d

M_s = Sloughed biomass from the rotating discs into the mixed liquor, grams/d

Figure 37. Flow diagram for the first stages of the RBC units.

A mass balance of the biomass in the mixed liquor tank, complete mix reactor, results in the equation:

$$V_1 \frac{dX_1}{dt} = QX_0 + M_s - QX_1 - M_i \qquad (35)$$

If biomass reactions in the mixed liquor are neglected. In applying Equation 35 to the first RBC stage, the influent biomass, X_0, was assumed to be zero.

A mass balance of the substrate in the mixed liquor tank yields

$$V_1 \frac{dS_1}{dt} = QS_0 + L_s - QS_1 - L_i \qquad (36)$$

With steady-state conditions, the derivatives with respect to time equal zero. Equations 35 and 36 then can be written:

$$M_S - M_i = QX_1 \qquad (37)$$

$$L_i - L_S = QS_0 - QS_1 \qquad (38)$$

Substitution of Equations 37 and 38 into Equation 34 gives:

$$YQ(S_0 - S_1) - QX_1 - k_d A_1 \overline{X}_1 = 0 \qquad (39)$$

Yield and decay rate determinations

Equation 39 can be rearranged as follows:

$$\frac{QX_1}{A_1\overline{X}_1} = YQ \frac{(S_0 - S_1)}{A_1\overline{X}_1} - k_d \qquad (40)$$

The steady-state data from the four experimental units were used to estimate Y and k_d by linear regression using Equation 40. Table C-1 (Appendix C) summarizes the values used in the variables for each temperature, and Table 32 summarizes the results. When calculating Y and k_d at 5°C, the data from unit A were excluded. Its incorporation resulted in a regression coefficient of 0.75 in contrast to a value of 0.998 without unit A data. Justification for excluding unit A data was based on the forced change in the experimental procedure where the influent wastewater ratio was changed after an attached growth was established. Although there were no significant changes with time in unit A data during a period extending over more than one month, the low temperature conditions may have delayed the response in the production of attached growth and the sloughing

rates. Figure 38 shows the regression lines for the data presented in Table C-1.

T-tests indicated that the difference between the yield coefficients at 5°C and 15°C was significant at the 40 percent level, and the difference between the 20°C and 15°C yield coefficients was significant at the 30 percent level. The yield coefficients at 5°C and 20°C did not differ at the 50 percent level. Although a high level of significance was not observed, there was an indication that the optimum yield coefficient occurred at 15°C.

The difference in decay coefficients at 5°C and 15°C was significant at 30 percent, and at 5°C and 20°C at 20 percent. The decay rates between 15°C and 20°C did not differ at the 50 percent level.

Table 32. Summary of the results of the linear regression analyses of the data used to calculate yield coefficients and decay rates.

Parameter	5°C	15°C	20°C
Yield coefficient, mg VS/mg COD	0.66	0.80	0.63
Decay rate, 1/d	0.07	0.22	0.26
Regression coefficient	0.998[a]	0.934	0.949
Significance level, %	1	10	5

[a]Based on three units; B, C, D.

The values of the yield coefficients are presented in Table 32 and also shown in Figure 39. The maximum yield of attached growth occurred at 15°C. Muck and Grady (1974), using an activated sludge mixed culture, observed an optimum in the yield coefficient at 20°C. Temperatures of 10°C, 20°C, 30°C, and 40°C were investigated. The yield coefficients of these temperatures were 0.469, 0.598, 0.515, and 0.445, respectively. Clark et al. (1978) reported an apparent yield coefficient of 0.96 based on BOD_5. The temperature values were not reported. The relationship between temperature and decay rates has been expressed by both of the following equations:

$$(k_d)_T = (k_d)_{20}\,\theta_d^{T-20} \tag{41}$$

$$k_d = A_r e^{-E_a/RT_a} \tag{42}$$

where

θ_d = temperature coefficient for decay rate

T = temperature, °C

A_r = frequency factor, 1/d

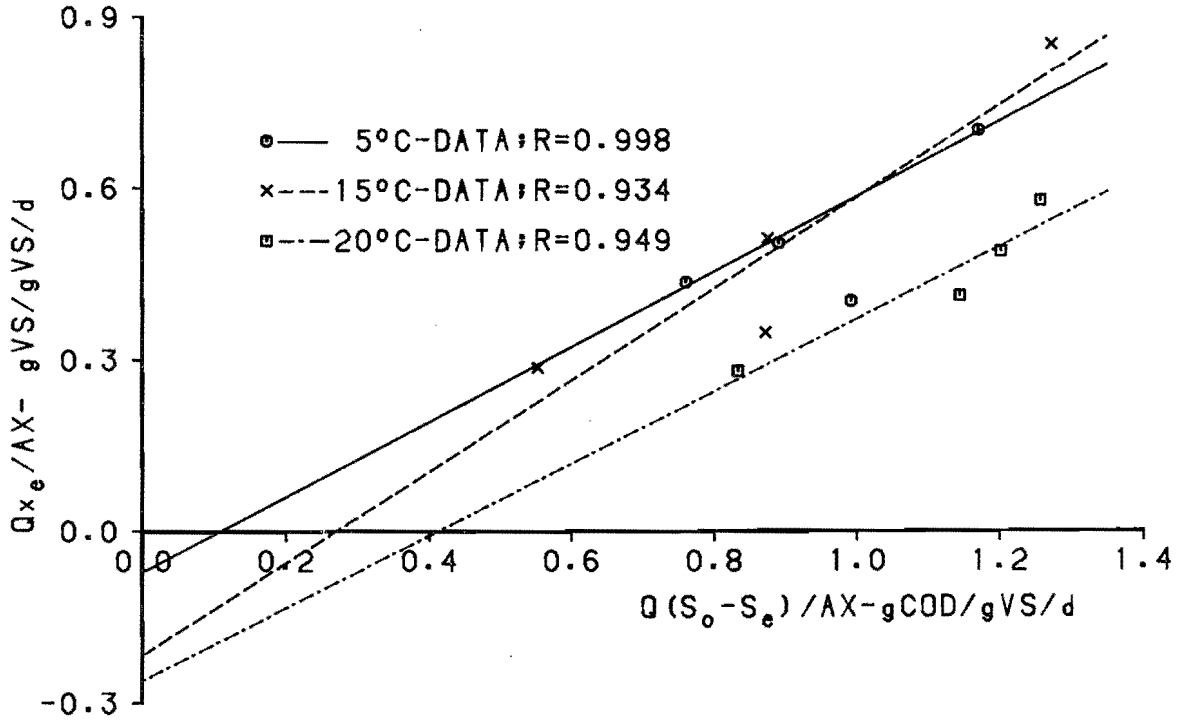


Figure 38. Results of regression analyses used to determine the yield coefficient and the decay rates.

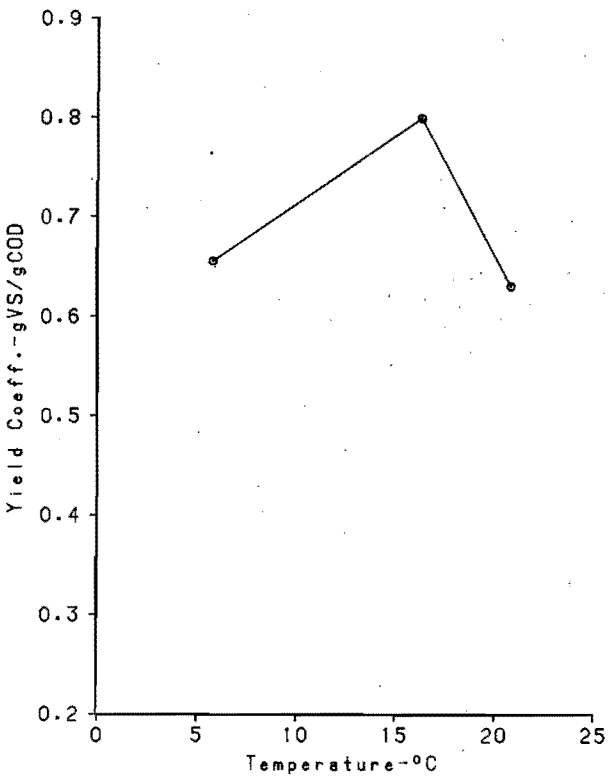


Figure 39. Relationship between the yield coefficient and temperature.

E_a = activation energy, cal/mole

T_a = absolute temperature, °K

k = universal gas constant = 1.987 cal/mole-°K

Equations 41 and 42 can be rearranged as follows:

$$\ln(k_d)_T = \ln\left[\frac{(k_d)_{20}}{\theta_d^{20}}\right] + T \ln \theta_d \qquad (43)$$

$$\ln k_d = \ln A_r - \frac{E_a}{R}\frac{1}{T_a} \qquad (44)$$

With the equations in this form, linear regression can be used to estimate the variables with the results shown in Table 33. Estimates were made assuming temperatures of 5.8°C, 16.3°C, and 20.8°C, which were the mean temperatures for the first stages. The mean temperature of 5.8°C was for units B, C, and D. The temperatures of 16.3°C and 20.8°C were based on all of the data from the four units. Figure 40 shows the relationship between temperature and decay rate values shown in Table 32. Muck and Grady (1979), using an activated sludge mixed culture, found the activation energy to be 13,900 cal/mole and the frequency factor value to be 2.031 x 10^8/hr or 0.487 x 10^{10}/d. These results compare favorably with Table 33. Unfortunately, data for other RBC units were not available for comparison.

Table 33. Temperature dependence of decay rate.

Parameter	Equation	
	$(k_d)_T = (k_d)_{20}\ \theta_d^{T-20}$	$k_d = A_1 e^{-E_a/RT_a}$
Correlation Coefficient	0.989	-0.989
k_{d20}, 1/d	0.27	
θ_d	1.09	
A_1 1/d		1.788×10^{10}
E_a, cal/mole		14,500

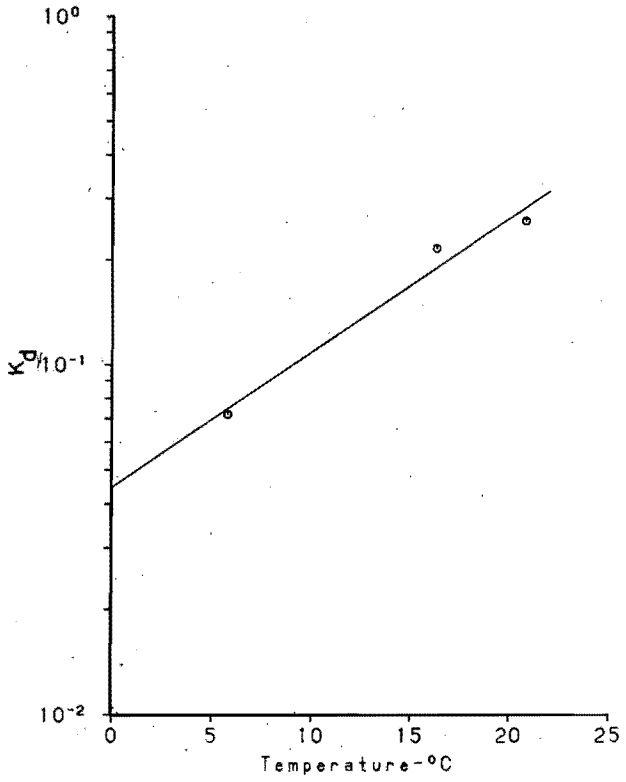


Figure 40. The relationship between the decay rate and temperature.

Carbonaceous Substrate Removal

General assumptions

In modeling carbonaceous substrate removal, it was assumed that the substrate is consumed only by the attached growth (substrate removal takes place only within the attached growth phase). For the first stage, the available substrate in the influent was the total substrate (expressed as total COD), and the unreacted substrate remaining after the first stage treatment was the filtered substrate (expressed as filtered COD) in the mixed liquor. For the second, third, and fourth stages the available influent substrate was the filtered COD from the preceding stage. The particulate material in the influent to these stages was assumed to be the sloughed biomass from the preceding stages.

Kinetic model for first stage carbonaceous substrate removal

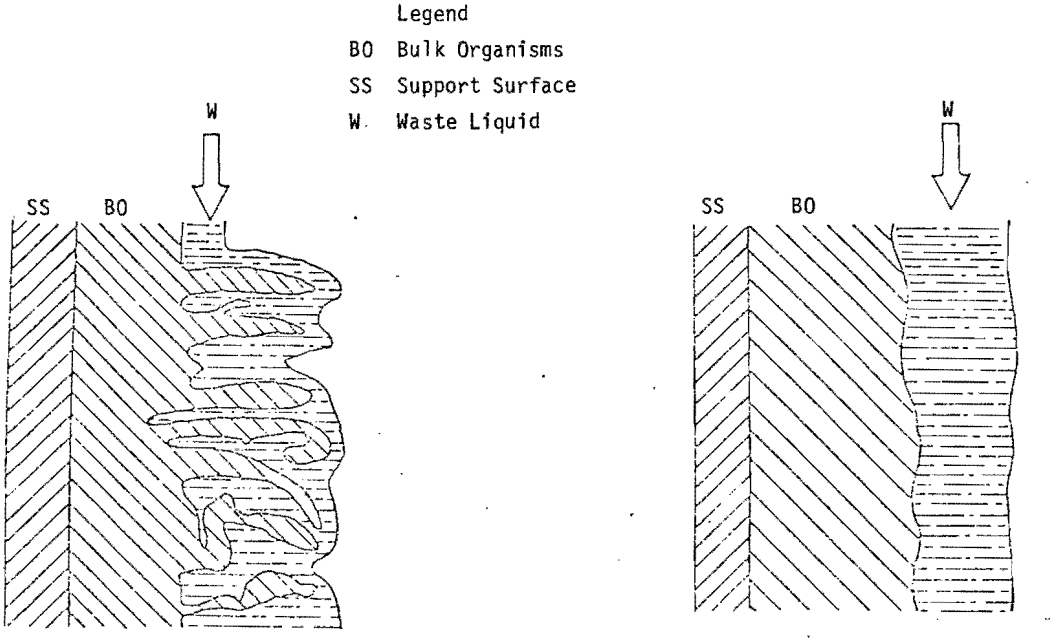
The experimental portion of this study showed that substrate removal in the first stage of an RBC unit could be described by substrate limiting kinetics; therefore, zero order or first order reaction kinetics were not used in the mathematical model development.

Either a pseudo homogeneous film model or a heterogeneous film model (Figure 41) can be used to represent carbonaceous substrate removal. The spongy appearance of the attached growth in the first stage of an RBC unit suggests a pseudo homogeneous film. Several approaches to describing the associated substrate removal kinetics were evaluated. A summary of them is presented in the following paragraphs.

a) One assumption is that the total attached biomass can be used to describe the active biomass in a Monod (hyperbolic relationship) kinetic model. The quantity of total attached biomass is dependent upon the organic loading rates and temperature; therefore, it is logical that the influence of organic loading rate and temperature on the attached biomass be incorporated into the model rather than use a constant mass of attached growth. Some portions of the biofilm experience substrate limitations and/or oxygen limitations to growth; however, at steady-state conditions it is logical to assume that the ratio of the active biomass to total biomass will be constant. As a result, the reaction rate constant will incorporate this ratio.

b) A constant active biomass layer can be applied to models in situations where substrate diffusion limitations control. Mathematical models developed for pseudo homogeneous films, including simultaneous reactions and substrate diffusion (Schroeder 1977, Friedman et al. 1976), may adequately describe the carbonaceous substrate removal kinetics. The reaction rate constants in this type model include a factor for the quantity of active biomass and the mass transport properties within the biofilm (Schroeder 1977).

c) A constant active biomass layer would be applicable in models describing systems where the total amount of attached biomass is governed by the physical conditions, such as mixing conditions. A mathematical model such as the one developed by Kornegay and Andrews (1968a) might adequately describe the carbonaceous substrate removal kinetics under physically limiting conditions. As has been shown in this study, physical conditions normally employed in RBC units do not seem to govern the amount of attached



A. Pseudo-Homogeneous Film B. Heterogeneous Film

Figure 41. Schematic presentation of pseudo-homogeneous film and heterogeneous film (Grieves 1972).

growth. Moreover, the similarity in the quantity of attached biomass per unit area observed in this study to those in other studies with different mixing speeds (Clark et al. 1978; Ouyang 1980; Mikula 1979) emphasizes that the physical conditions employed in this study were not unique.

Following the nomenclature employed in Figure 37, mass balances of the substrate around the attached growth phase and around the mixed liquor tank are represented by Equations 45 and 46, respectively.

$$A_1\frac{dS^*}{dt} = L_i - L_s - rA_1 \quad (45)$$

$$V_1\frac{dS_1}{dt} = QS_0 + L_s - QS_1 - L_i \quad (46)$$

where

A_1 = total available surface area for attached growth, m^2

S^* = substrate concentration within the attached growth, grams/m^2

L_i = substrate adsorbed and absorbed from the mixed liquor, grams/d

L_s = substrate released from the rotating attached growth into the mixed liquor, grams/d

r = reaction rate of carbonaceous substrate removal, grams/m^2/d

V_1 = mixed liquor volume in the tank, m^3

Q = influent wastewater flow rate, m^3/d

S_0 = influent substrate concentration, mg/l

S_1 = effluent substrate concentration, mg/l

For steady-state conditions the derivatives with time in Equations 45 and 46 equal zero. Adding these two equations represents steady-state with the relationship:

$$QS_0 - QS_1 - rA_1 = 0 \quad (47)$$

The three alternatives discussed above for quantifying the active biomass are applied and compared below.

a) Applying the total attached biomass with Monod growth kinetics gives:

$$r = \frac{k\,\bar{X}_1\,S_1}{K_s + S_1} \quad (48)$$

$$k = \frac{\hat{\mu}}{Y} \quad (49)$$

where

k = maximum reaction rate, 1/d

$\hat{\mu}$ = maximum specific growth rate, 1/d

Y = yield coefficient

$\bar{X}_1$ = total attached biomass, gVS/m^2

K_s = half-saturation constant, mg/l (in this study COD)

S_1 = substrate concentration in the tank, mg/l (in this study COD)

b) Applying a simultaneous substrate consumption and diffusion model (Schroeder 1977; Friedman et al. 1976) with constant active biomass gives:

$$r = \frac{k_m\,S_1^{\,2}}{K_m + S_1} \quad (50)$$

where

k_m = maximum reaction rate (grams/m^2)/(mg/l·d)

K_m = constant, mg/l

c) Applying a constant active biomass model (Kornegay and Andrews 1968a) with a Monod growth kinetics model gives:

$$r = \frac{k_a\,S_1}{K_s + S_1} \quad (51)$$

where

k_a = maximum reaction rate, grams/m^2/d

K_s = half-saturation constant, mg/l

Estimation of first stage carbonaceous substrate removal kinetic constants

Equations 48, 50, and 51, representing three different approaches to substrate removal kinetics were substituted into Equation 47 and rearranged in linear form. The resulting equations are, respectively:

$$[Q(S_0-S_1)/A_1\bar{X}_1]^{-1} = \frac{K_s}{k}\,\frac{1}{S_1} + \frac{1}{k} \quad (52)$$

$$[Q(S_0/S_1-1)/A_1]^{-1} = \frac{K_m}{k_m}\,\frac{1}{S_1} + \frac{1}{k_m} \quad (53)$$

$$[Q(S_0-S_1)/A_1]^{-1} = \frac{K_s}{k_a}\,\frac{1}{S_1} + \frac{1}{k_a} \quad (54)$$

The experimental steady-state data collected at each temperature were used in the above equations to estimate the kinetic constants. The data are summarized in Table C-2. Only Equation 52, which

was derived from Equation 48, yielded reasonable values for the kinetic constants (K_s or K_m) for all three temperatures (Table 34). The mass transport model (Equation 53) produced reasonable values only with the data obtained at 5°C and 15°C. At these temperatures the values for K_m were 20.8 mg/l and 42.5 mg/l, which are close to those obtained by Friedman et al. (1976) and Friedman et al. (1979). At 20°C the mass transport model resulted in a negative value for K_m. The negative value may have occurred, among other reasons, because at high temperatures the kinetics are described by substrate limiting conditions and not diffusion. Applying fixed biomass quantities with Monod growth kinetics as expressed in Equation 54 resulted in negative values for both the kinetic constant K_s and the reaction rate k_a at temperatures of 5°C and 15°C. At 20°C the reaction rate was determined to be 174 grams/m^2/d and the half-saturation constant was 111.5 mg/l (Table 34). Kornegay and Andrews (1968a) reported K_s value of 121 mg/l glucose in experiments conducted at 25°C.

The 5°C data were regressed for all the units and for the three units B, C, D. As mentioned earlier, using the data from unit A at 5°C for yield and decay determination resulted in low correlation coefficients. As indicated in Table C-3, there were no significant differences between the kinetic constants obtained with the two sets of data, although there was a decrease in the correlation coefficients when unit A was included. This decline in the goodness of fit might indicate that in unit A (at 5°C) the attached biomass quantity and/or its sloughing rate did not achieve steady state.

When data from unit A at 20°C were used, the regression analyses for Equations 52, 53, and 54 resulted in very low correlation coefficients or negative kinetic values (Table C-3). The relatively low performance of unit A at 20°C when compared with the other units is shown in Figures 5 and 20 where filtered COD and filtered TOC removals were low. This low removal may have been caused by the release of unbiodegradable material through cell lysing. Consequently unit A data were omitted in evaluating the kinetic constants.

Based upon the above analyses, Equation 52 seems to be the most reliable of the three in describing first stage carbonaceous substrate removal. Rearranging the terms,

$$Q(S_0-S_1) = A_1 \frac{k \bar{X}_1 S_1}{K_s + S_1} \qquad (55)$$

Figure 42 shows the curves calculated using Equation 55 and the determined constants along with the measured data. The constant, k, was defined by Equation 49 as $\hat{\mu}/Y$. Kinetic constants obtained for the first stages of the RBC units are summarized in Table 35.

Table 34. Summary of the kinetic constants for carbonaceous substrate removal in the first stages.[a]

Equation	Temperature, °C								
	5			15			20		
	Correlation Coefficient	k, k_m, k_a	K_s, K_m	Correlation Coefficient	k, k_m, k_a	K_s, K_m	Correlation Coefficient	k, k_m, k_a	K_s, K_m
$Q(S_0-S_1) = A_1 \frac{k \bar{X}_1 S_1}{K_s+S_1}$ (52)	0.965	2.85	61.6	0.950	7.76	262.2	0.999	9.44	276.4
$Q(S_0-S_1) = A_1 \frac{k_m S_1^2}{K_m+S_1}$ (53)	0.893	1.12	20.8	0.886	1.80	42.5	-0.808	0.98	-5.8
$Q(S_0-S_1) = A_1 \frac{k_a S_1}{K_s+S_1}$ (54)	0.986	-41.0	-93.6	0.965	-36.3	-81.6	0.983	174	111.5

[a] For 5°C and 20°C only the data from units B, C and D were used, for 15°C the data from units A, B, C and D were used.

Q = Flow rate, m^3/d
S_0, S_1 = Influent and Effluent COD, mg/ℓ
A_1 = Total surface area/stage, m^2
k = Maximum reaction rate, 1/d
k_m = Maximum reaction rate, grams/m^2/mg/ℓ/d
k_a = Maximum reaction rate, grams/m^2/d
$\bar{X}_1$ = Total attached biomass in first stage, gVS/m^2
K_s = Half saturation constant, mg/ℓ COD
K_m = Constant, mg/ℓ COD

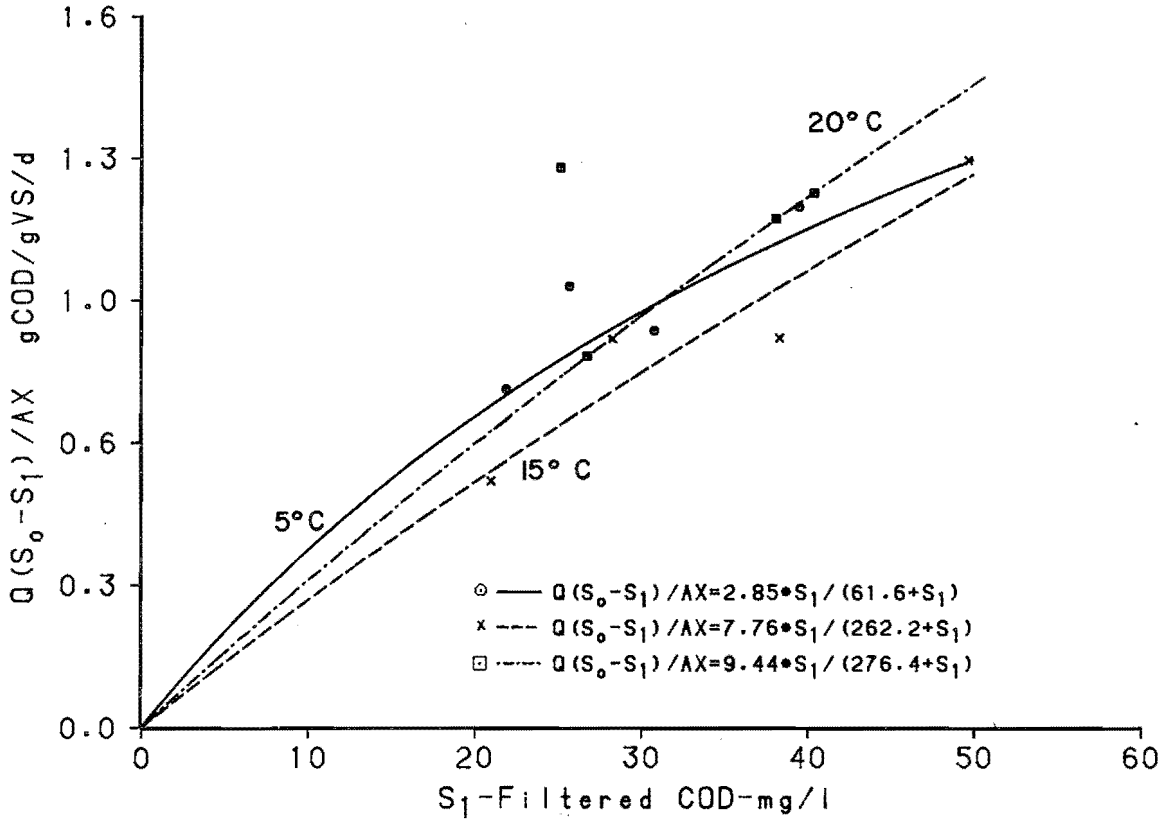


Figure 42. The relationship between the calculated and actual specific substrate removal rates and the substrate concentrations.

Table 35. Summary of kinetic constants obtained for the first stages of the RBC units.

Parameter	Temperature, °C		
	5	15	20
Y = yield coefficient, mgVSS/mg COD	0.66	0.80	0.63
$\hat{\mu}$ = maximum specific growth, 1/d	1.87	6.20	5.95
k = maximum reaction rate, 1/d	2.85	7.76	9.44
k_d = decay rate, 1/d	0.07	0.22	0.26
K_s = half saturation constant, mg/ℓ COD	61.6	262.2	276.4

The maximum specific growth rates shown in Figure 43 indicate an optimum temperature of about 15°C, as was indicated also with the yield coefficients. Kornegay and Andrews (1968) reported $\hat{\mu}$ values of 6.72/d for their experiments at 25°C with the rotating drum using glucose as substrate. Clark et al. (1978) reported maximum specific growth rates of 4.4/d in the first stage and a half-saturation constant of 431 mg/l of soluble BOD_5. The temperature values were not reported. Using the total biomass as done in this study, rather than the 70 percent used as the active portion by Clark et al. (1978), yields a maximum specific growth rate of 6.3/d (4.4/0.7 = 6.286).

The maximum reaction rate (k) can be related to temperature using functions similar to those used with the decay rate. The Arrhenius equation relates the reaction rate and temperature as follows:

$$k_T = A_r e^{-Ea/RT_a}, \quad (56)$$

Alternatively, the conventional expression used for wastewater treatment processes is:

$$k_T = k_{20}\ \theta_s^{T-20} \quad (57)$$

where

k_T, k_{20} = maximum reaction rate at temperature T°C and 20°C, 1/d

A_r = frequency factor, 1/d

E_a = activation energy, cal/mole

T_a = absolute temperature, °K

T = temperature, °C

R = universal gas constant = 1.987 cal/mole-°K

θ_S = temperature coefficient for carbonaceous substrate removal

Table 36 summarizes the parameter estimates found linear regression after rearranging Equations 56 and 57 to the linear forms:

$$\ln k_T = \ln \left(\frac{k_{20}}{\theta_s^{20}}\right) + T \ln \theta_s \quad (58)$$

$$\ln k_T = \ln A_r - \frac{E_a}{R}\ \frac{1}{T_a} \quad (59)$$

The temperatures used in the regression analyses were 5.8°C, 16.3°C and 20.8°C, which were the mean temperatures for the first stages of the RBC units. The temperature factor of 1.09 obtained with the conventional equation was similar to the typical value of 1.08 for the trickling filter process (Metcalf and Eddy, Inc. 1979). An activation energy of 13,900 cal/mole was reported by Murphy et al. (1977) for nitrification reaction rates.

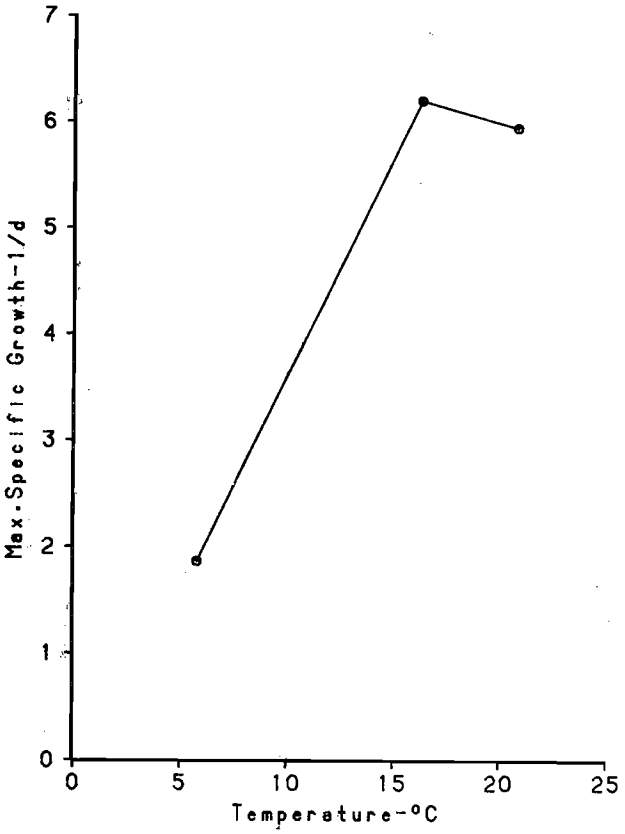


Figure 43. Relationship between the maximum specific growth and temperature.

Figures 40 and 44 show the temperature dependence of decay rate and maximum reaction rate, respectively. The two patterns are similar. For both processes, the temperature factor was 1.09. The relationships in

Table 36. Relationships between temperature and the maximum reaction rate.

Parameter	Equation $k_T = k_{20}\theta_s^{T-20}$	Equation $k_T = A_r e^{-E_a/RT_a}$
Correlation Coefficient	0.990	-0.990
k_{20}, 1/d	9.50	
θ_s	1.09	
A_r, 1/d		9.934×10^{10}
E_a, cal/mole		13,400

Figures 40 and 44 also show that the increase in these rates between 5°C and 15°C is greater than the increase that occurs between 15°C and 20°C. This relationship may explain why there was no significant increase in BOD_5 removal in full scale RBC units beyond a temperature of 12.7°C (55°F) (Antonie 1976). The experiments described in this report show that a significant increase in the substrate removal occurs in the first stages as temperatures pass 15°C. In terms of overall removal for the four stages, however, the differences between 15°C and 20°C were less.

The results presented in Table 35 and plotted in Figure 45 also show that there was an increase in the half-saturation constant as the temperature was increased from 5°C to 15°C. Beyond 15°C there were no significant changes in the values obtained for the half-saturation constant. No reports could be found in the literature on RBC treatment on the effect of temperature on the half-saturation constant (K_s). Muck and Grady (1979) reported a slight decrease in K_s in an activated sludge system as the temperature was changed from 10°C to 20°C and a significant increase when the temperature was increased from 20°C to 30°C.

The differences between the maximum reaction rates (k_T) at 5°C and those at 15°C and 20°C are statistically significant at a significance level of 30 percent, and 20 percent, respectively. The differences between the k_T values at 15°C and 20°C are not significant at a level of 50 percent. The significance test (t-test) for differences between the slopes of (K_s/k_T) of Equation 52 indicated that the slopes do not differ at a significance level of 50 percent. This lack of a difference may be attributable to the narrow interval of S_1 obtained in the experiments.

The kinetic model for carbonaceous substrate removal (Equation 55) incorporates the total quantity of attached biomass in the first stages. The attached biomass on an RBC unit cannot be controlled economically as can be done in an activated sludge culture; however, the results of this study indicated that the

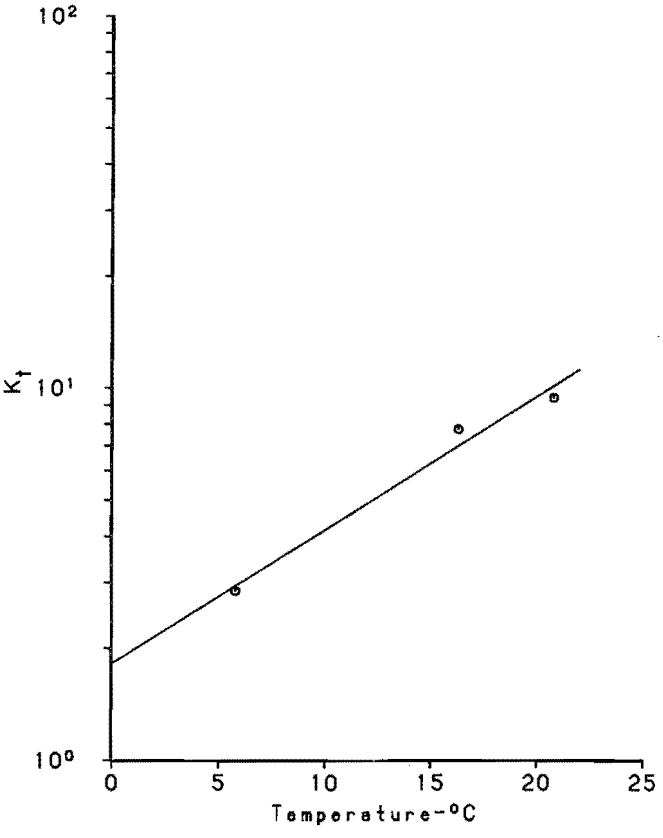


Figure 44. The relationship between the maximum reaction rate and temperature.

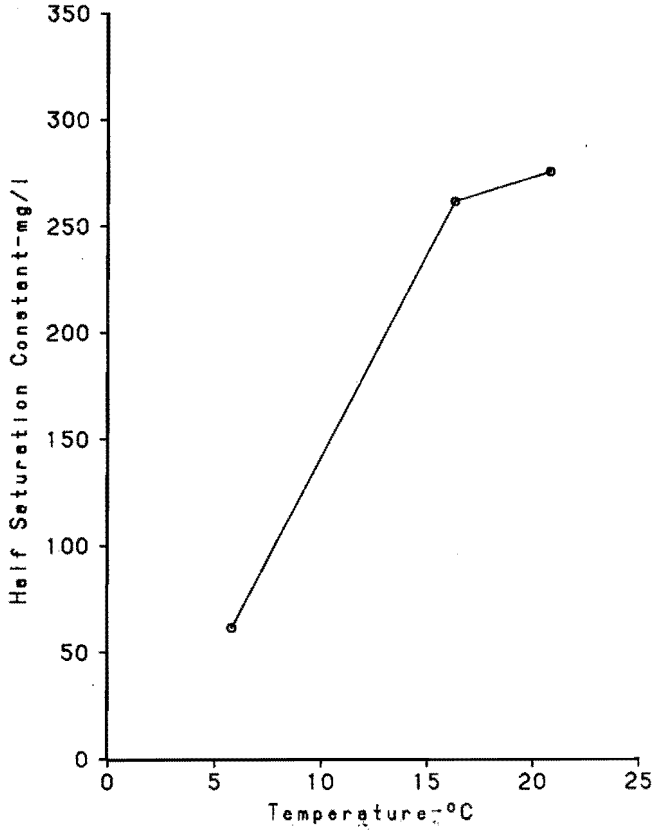


Figure 45. The relationship between the half-saturation constant and temperature.

quantity of attached biomass is predictable. A saturation type relationship that can be used for a given temperature to predict the quantity of attached biomass is:

$$\bar{X}_1 = \frac{k_X M_1}{K_X + M_1} \quad (60)$$

where

$\bar{X}_1$ = the quantity of attached biomass in the first stage per unit surface area, gVS/m^2

M_1 = organic load per first stage surface area, gCOD/m^2/d

k_X = constant, gVS/m^2 K_X = constant, gCOD/m^2/d

For linear regression, Equation 60 was converted into the linear form:

$$\frac{1}{\bar{X}_1} = \frac{K_X}{k_X}\frac{1}{M_1} + \frac{1}{k_X} \quad (61)$$

The regressions for estimating the two constants gave correlation coefficients of 0.919 and 0.972 (significance level = 5 and 1 percent, respectively) for 5°C and 15°C data, respectively. Estimation with the 20°C data resulted in a negative value for k_X. Omitting unit A data and assuming the same K_X value obtained at 15°C, k_X was reestimated by minimizing the sum of the squares of the deviations between the predicted and the observed values (Table 37). Figure 46 presents the curves based on the estimated constants along with the measured data.

The values of k_X were related to the temperature using Equation 57. The correlation coefficient was 0.986, and the relationship is:

$$(k_X)_T = 56.9(1.015)^{T-20} \quad (62)$$

Table 37. Summary of the first stage attached biomass constants.

Constant	Mean Temperature, °C		
	5.9	16.3	20.8
k_X	46.15	52.54	58.50
K_X	31.07	23.77	23.77

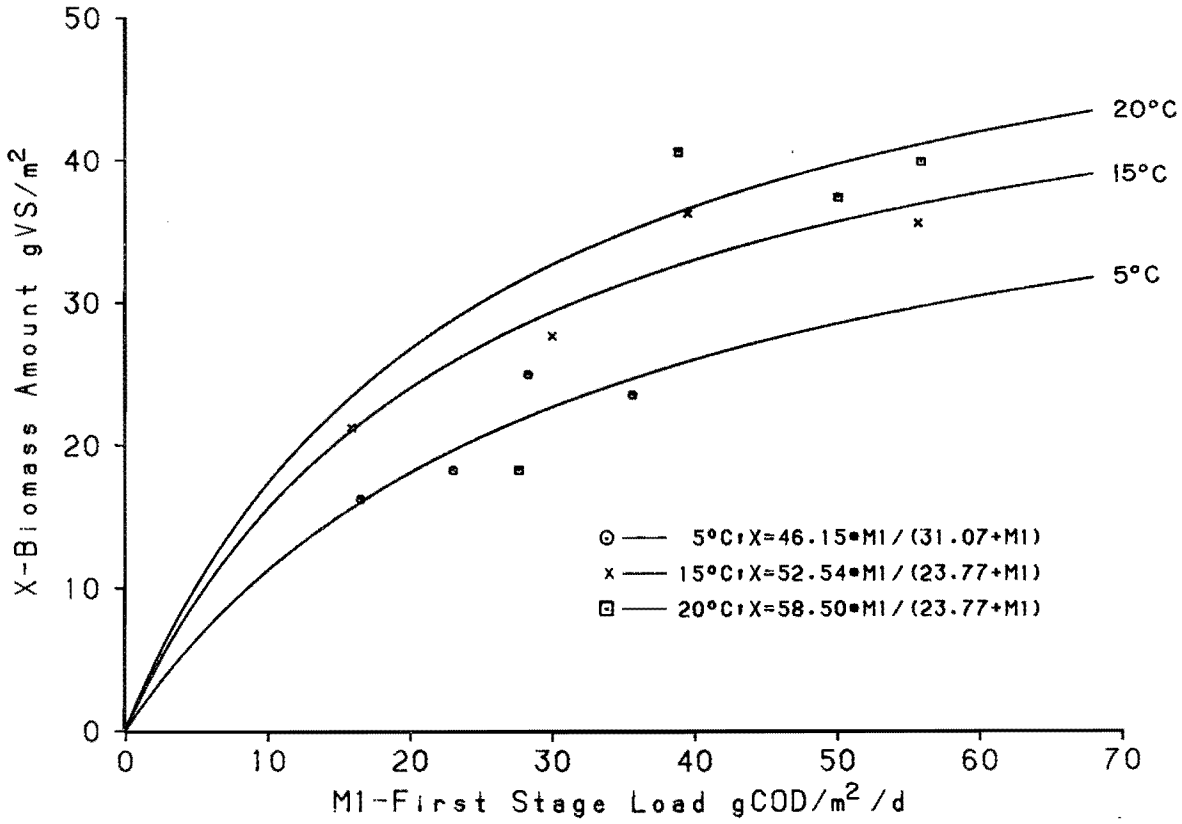


Figure 46. Relationship between the quantity of attached biomass produced in the first stages and the organic loading rate at various temperatures.

where

T = °C

$(k_x)_T$ = gVS/m²

Figure 47 shows the relationship plotted. The values for temperatures between 5°C and 15°C can be estimated by simple proportion because the difference between the values is not large.

Kinetic model for second through fourth stage carbonaceous substrate removal

The experimental work showed that the major part of the carbonaceous substrate removal occurred in the first stage. The only significant substrate removal (expressed as filtered COD) in the succeeding stages occurred in the highly loaded units C and D and then only in the second stage. As a result, it was not possible to apply a rational kinetics model to each stage and temperature and determine the kinetic coefficients because of the limited number of degrees of freedom.

Substitution of an empirical model was attempted. For this model, the second through fourth stages were considered as one complete mix reactor because the concentration of filtered COD in the effluent did not follow a trend. The mean filtered COD concentrations from the last three stages were assumed to represent the substrate concentration in the hypothetical reactor. The models applied with the data summarized in Table C-4 were:

$$Q(S_1-\bar{S}) = \sum_{i=1}^{3} A_i \, (k_L)_{20} \, \theta_L^{T-20} \, S_1^n \qquad (63)$$

$$Q(S_1-\bar{S}) = \sum_{i=1}^{3} A_i \, (k_L)_{20} \, \theta_L^{T-20} \, (\bar{S})^n \qquad (64)$$

where

Q = influent flow rate, m^3/d

S_1 = first stage substrate concentration mg/l

$\bar{S}$ = the mean substrate concentration in the second through the fourth stages, mg/l

A_i = total available surface area/stage, m^2/stage

$(k_L)_{20}$ = reaction rate at 20°C, $g/m^2/d$

θ_L = temperature factor

T = temperature, °C

n = apparent reaction order

Multiple regression analyses were used to estimate the three parameters after rearranging Equations 63 and 64, respectively, to:

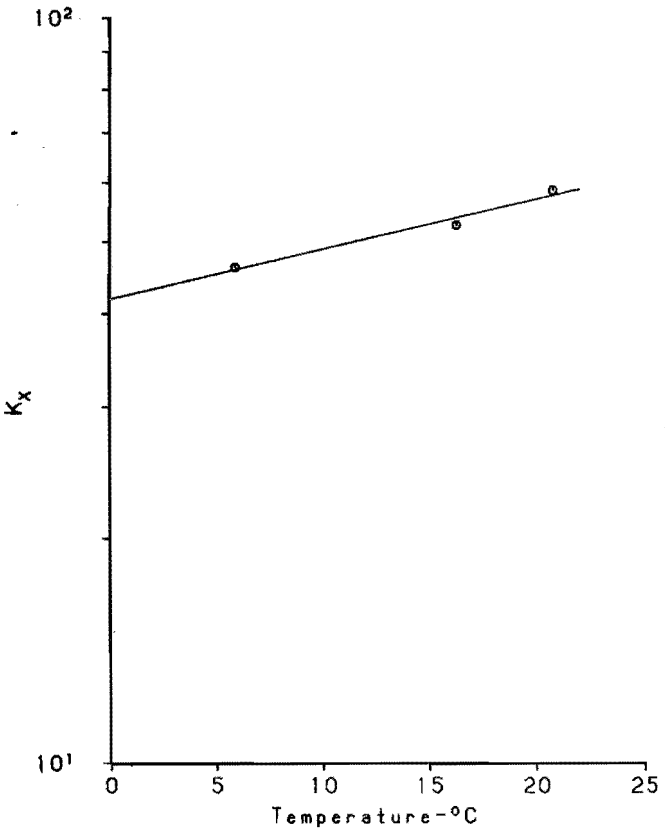


Figure 47. Relationship between k_x and temperature.

$$\ln\left[\frac{Q(S_1-\bar{S})}{\sum_{i=1}^{3} A_i}\right] = \ln(k_L)_{20} + (T-20)\ln\theta_L + n\ln S_1 \qquad (65)$$

$$\ln\left[\frac{Q(S_1-\bar{S})}{\sum_{i=1}^{3} A_i}\right] = \ln(k_L)_{20} + (T-20)\ln\theta_L + n\ln\bar{S} \qquad (66)$$

Table 38 summarizes the results. These results show that the temperature factor for carbonaceous substrate removal in the second through fourth stages was approximately the same as the temperature factor in the first stages (Table 36). This suggests that a single temperature factor could be applied for carbonaceous substrate removal kinetics through all the stages in the RBC process.

The reaction order for Equation 64 was 0.688 as shown in Table 38. Rittmann and McCarty (1978, 1980) showed that substrate removal kinetics could be expressed as variable order kinetics with values ranging between 0.5 to 1.0 for heterogeneous models of attached growth. As discussed previously, in the last stages of RBC units heterogeneous models of attached growth are justified. Consequently, the reaction order of 0.688 is reasonable to express the

Table 38. Summary of results of the regression analyses for carbonaceous substrate removal in the second through fourth stages of RBC units.

Equation Number	N*	R^2	Reaction Rate-$(k_L)_{20}$ g/m²/d	Reaction Order-n	Temperature factor-θ_L
65	7	0.986	0.0444	0.763	1.11
66	7	0.984	0.0715	0.688	1.12

*N = number of determinations

carbonaceous substrate removal in the last stages of RBC process treating domestic wastewater.

For simulation purposes, Equation 63 is more convenient to use and will yield essentially the same results.

$$\frac{Q(S_1-\bar{S})}{\sum_{i=2}^{4} A_i} = 0.0444\ (1.11)^{T-20}\ S_1^{0.763} \qquad (63)$$

Equation 63 can be used to determine $\bar{S}$ but not the final effluent filtered COD. With low organic loading rates, such as those used with the experimental units, applied to an RBC system, significant differences between S_4 and $\bar{S}$ are not likely (S_4 = final effluent filtered COD concentration, $\bar{S}$ = mean value of the filtered COD concentration in the last three stages). For higher organic loading rates where values of S_1 may approach 100 mg/l COD, then the model for the first stage can be applied to the second stage. If such were the case, the denominator on the left side of Equation 63 would be:

$$\frac{Q(S_2-\bar{S})}{\sum_{i=3}^{4} A_i} = 0.0444\ (1.11)^{T-20}\ S_2^{0.763} \qquad (67)$$

Sloughed Biomass Stabilization

General assumptions and difficulties

Sloughed biomass stabilizes as it passes through the stages. Part of the sloughed biomass reattaches on the following discs. The reattached biomass was assumed to be in the endogenous phase of stored substrate consumption. Adsorbed substrate on the biomass continued to be available food for growth and maintenance as did the filtered substrate in the mixed liquor. It was assumed that the yield coefficient for filtered substrate consumption was the same for all stages, and the value determined for the first stage was used.

The major problem in estimating the kinetic coefficients of the biomass stabilization process was the relative instability of the attached biomass in the later stages. The instability may have been caused by low organic loading rates and the dynamic nature of the sloughing process. The steady-state assumption used to develop a mathematical model for biomass stabilization was at best a gross approximation. To avoid consequent fluctuations in estimating the coefficients, the last three stages were again considered as one complete mix reactor. The mean concentrations of particulate and filtered COD from the three stages were assumed to represent the complete mix reactor concentrations. The total attached biomass for the combined reactor was calculated by summing the mass of attached biomass for the three stages.

Kinetic model development for sloughed biomass stabilization

Figure 48 shows a flow diagram of the stabilization process for sloughed biomass for use in writing mass balance relationships. Filtered substrate consumption yields particulate material. Using the ratios of particulate COD to VSS of slough biomass, the yield of particulate material in terms of COD is

$$\sum_{i=2}^{4} A_i \left(\frac{dM}{dt}\right)_{production} = Y \cdot a\ (L_i - L_S) \qquad (68)$$

in which

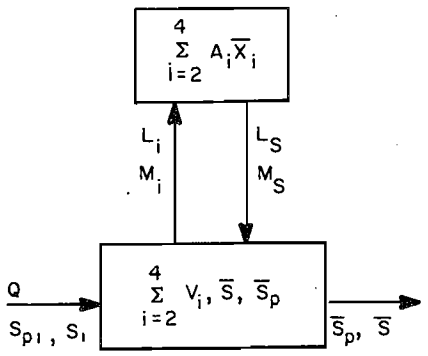
M = particulate material on the discs, gCOD/m²

Y = yield coefficient

a = grams of particulate COD/gVSS

A mass balance of the particulate material expressed as COD around the attached growth gives:

$$\sum_{i=2}^{4} A_i \frac{dM}{dt} = M_i - M_S + \sum_{i=2}^{4} A_i \left(\frac{dM}{dt}\right)_{production} - \sum_{i=2}^{4} A_i r_{st} \qquad (69)$$



Q = Influent flow, m^3/d

$S_{p1},\overline{S}_p$ = Influent and effluent particulate substrate, mg/l COD

$S_1,\overline{S}$ = Influent and effluent filtered substrate, mg/l COD

L_i,L_S = Filtered substrate adsorption and

M_i,M_S = Particulate substrate adsorption and release, g/d

A_i = Total available surface area per stage, m^2

$\overline{X}_i$ = Attached growth at stage i, g/m^2

V_i = Mixed liquor volume per stage, m^3

Figure 48. Flow diagram for the sloughed biomass stabilization process.

in which

r_{st} = reaction rate for particulate COD removal, $gCOD/m^2/d$

The reaction rate, r, can be expressed in a relationship similar to the one used to describe the decay rates.

$$r_{st} = k_{st} \sum_{i=2}^{4} \overline{X}_i \qquad (70)$$

in which

k_{st} = stabilization constant, gCOD/g attached growth/d

Substituting the value of r_{st} from Equation 70 and the particulate material production from Equation 68 into Equation 69 gives:

$$\sum_{i=2}^{4} A_i \frac{dM}{dt} = M_i - M_S + Ya\,(L_i-L_S) - \sum_{i=2}^{4} A_i\overline{X}_i k_{st} \qquad (71)$$

A mass balance of the filtered substrate around the mixed liquor, assuming that all substrate removal occurs on the attached biomass, gives:

$$\sum_{i=2}^{4} V_i \frac{d\overline{S}}{dt} = Q\,S_1 - Q\overline{S} + L_S - L_i \qquad (72)$$

A mass balance of the particulate COD around the mixed liquor gives:

$$\sum_{i=2}^{4} V_i \frac{d\overline{S}_p}{dt} = QS_{p1} - Q\overline{S}_p + M_S - M_i \qquad (73)$$

Under steady-state conditions, the derivatives with respect to time in the above two equations are equal to zero. Substitution of the term M_i-M_S from Equation 73 and the term L_i-L_S from Equation 72 into Equation 71 yields:

$$Q(S_{p1}-\overline{S}_p) + YaQ(S_1-\overline{S}) - \sum_{i=2}^{4} A_i\overline{X}_i k_{st} = 0 \qquad (74)$$

Estimation of sloughed biomass stabilization kinetic constants

Table 39 shows a summary of the data used to calculate the sloughed biomass stabilization constants. Only the data from the experiments in which particulate COD removals were measured were used in the calculations. The a (defined for Equation 68) values used in Equation 74 were 1.434, 1.079, and 1.366 (from Table 16) for the experimental data collected at 5°C, 15°C, and 20°C, respectively. Yield coefficients (Y) of 0.66 (5°C), 0.80 (15°C) and 0.63 (20°C) were used. Table 40 shows a summary of the values obtained using Equation 74 with the above referenced data to calculate k_{st}. The temperature factor, θ_{st}, was calculated by linearizing Equation 75 and using the mean values for each temperature and the values of k_{st} in Equation 76 to determine the best fit for the data.

$$k_{st_T} = (k_{st})_{20}\,\theta_{st}^{T-20} \qquad (75)$$

$$\ell n(k_{st})_T = \ell n(k_{st})_{20} + (T-20)\ell n\theta_{st} \qquad (76)$$

The following results were obtained:

correlation coefficient = 0.999;

$(k_{st})_{20}$ = 0.208 gCOD/g attached growth/d; and

θ_{st} = 1.037.

Table 39. Summary of the data used to calculate the stabilization constants for the sloughed biomass.

Temperature °C	Unit	Influent Flow, ℓ/d	Influent Particulate COD mg/ℓ	Influent Filtered COD mg/ℓ	Mixed Liquor Particulate COD, mg/ℓ	Mixed Liquor Filtered COD mg/ℓ	Three Stages Attached Biomass, grams
5	C	294	107.1	30.8	90.1	29.1	46.5
15	B	285.3	64.2	28.3	25.0	23.0	48.8
15	C	282.0	78.3	38.3	54.7	32.8	74.6
15	D	288.6	115.3	49.7	77.8	41.9	82.4
20	B	283.7	71.2	26.8	62.8	32.1	11.7
20	D	292.2	154.1	40.4	130.7	29.2	46.3

The stabilization constants are approximately equivalent to the decay values obtained in the first stages. Assuming that the attached biomass has the same ratio of particulate COD to VS (not measured) as the sloughed biomass, then the constant k_{st} can be expressed with dimensions of 1/unit of time.

$$k'_{st} = \frac{k_{st}}{a} \qquad (77)$$

where

k'_{st} = stabilization constant, 1/d

a = gCOD/gVSS

Substitution of the a values from Table 16 and the k_{st} values from Table 40 in Equation 77 results in k'_{st} values of 0.082/d, 0.160/d, and 0.152/d respectively for 5°C, 15°C, and 20°C.

The similarity between the stabilization constants and the decay rates (see Table 35) shows that the particulate COD, or the sloughed biomass while passing through the stages, was being stabilized. The temperature factor of 1.037 obtained in the regression analysis was lower than expected when compared with the other processes. An explanation may be that data from only one unit of the 5°C experiments were used in the calculations, and the results from this 5°C experiment appear to be slightly high when compared with the decay coefficients.

Table 40. Summary of stabilization constants calculated using Equation 74.

Temperature °C	Unit	k_{st} gCOD/gVS/d
4.6	C	0.118
14.6	B	0.256
14.5	C	0.107
15.1	D	0.155
20.2	B	0.204
19.7	D	0.209

Ammonia Nitrogen Removal

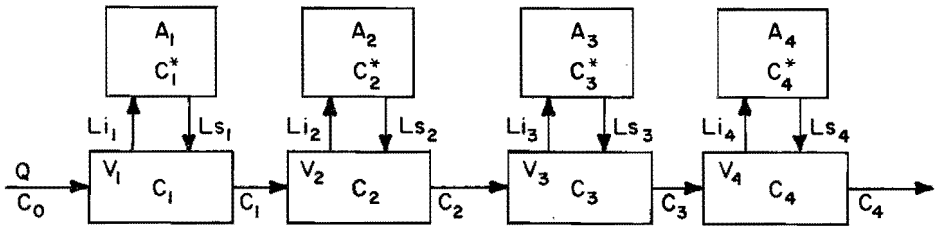
General assumptions

The results of the study showed that about 90 percent of the ammonia nitrogen removal or conversion in the RBC units was the result of nitrification. The kinetic model used for ammonia nitrogen removal assumed that this one process dominates in a mathematical model that incorporates nitrification along with other minor processes such as ammonia nitrogen stripping. It was assumed that the kinetic model for ammonia nitrogen removal was the same for all stages where nitrification was not inhibited.

Where substrate limiting conditions did not occur, ammonia nitrogen removal rates were the same in the second and third stages of the experimental RBC units. Based upon this observation, it was assumed that the kinetics were the same in all of the stages. Ammonia nitrogen removal rates were independent of the amount of attached growth at each temperature; therefore, the model was developed using a constant amount of active microorganisms to account for ammonia nitrogen consumption at a given temperature. The results of the experiments showed that the kinetics of ammonia nitrogen removal could be described by Monod growth kinetics. Since very little nitrification occurred at 5°C, only the results obtained at 15°C and 20°C were used in the development of the mathematical model to show the temperature dependence of the kinetic constants.

Kinetic model for ammonia nitrogen removal

A mathematical model for ammonia nitrogen removal is developed in the following paragraphs from data collected on the stages in which environmental conditions do not limit the growth of nitrifiers. Figure 49 shows a schematic flow diagram for a four stage RBC process in which nitrification is occurring and defines the symbols used in the ammonia nitrogen removal model development. It was assumed that each stage was a complete mix reactor and that consumption of ammonia nitrogen by the suspended growth was negligible.



A_i = Total available surface area per stage, m^2

V_i = Mixed liquor volume per stage, m^3

Q = Influent flow rate, m^3/d

C_0 = Influent ammonia nitrogen concentration, mg/l

C_i, i=1,4 = Mixed liquor and effluent ammonia nitrogen concentrations in stage i, mg/l

C_i^*, i=1,4 = Ammonia nitrogen concentration within the attached growth, g/m^2

L_{ii}, L_{Si}, i=1,4 = Ammonia nitrogen extraction rates from the mixed liquor, and release rates from the attached growth into the mixed liquor at stage i, g/d

Figure 49. Schematic flow diagram of a four stage RBC process.

A mass balance of the attached growth in the i stage where nitrifiers are growing gives

$$A_i \frac{dC_i^*}{dt} = L_{ii} - L_{Si} - A_i \frac{k_N C_i}{K_N + C_i} \quad (78)$$

A mass balance of the ammonia nitrogen in the mixed liquor of stage i gives

$$V_i \frac{dC_i}{dt} = Q\,C_{i-1} - Q\,C_i + L_{Si} - L_{ii} \quad (79)$$

At steady-state, the derivatives in Equations 78 and 79 equal zero; therefore, Equation 79 can be rewritten as:

$$L_{ii} - L_{Si} = Q\,C_{i-1} - Q\,C_i \quad (80)$$

Substitution of the left term of Equation 80 into Equation 78 yields:

$$Q\,C_{i-1} - Q\,C_i = A_i \frac{k_N C_i}{K_N + C_i} \quad (81)$$

in which

k_N = Ammonia nitrogen maximum reaction rate, $gN/m^2/d$

K_N = Ammonia nitrogen removal half-saturation constant, mg/l-N

Estimation of ammonia nitrogen removal kinetic constants

The Monod growth kinetics relationship of Equation 81 can be rewritten in linear form as:

$$\left[\frac{Q(C_{i-1} - C_i)}{A_i}\right]^{-1} = \frac{K_N}{k_N}\,\frac{1}{C_i} + \frac{1}{k_N} \quad (82)$$

To avoid distortion of the parameter estimates from large values of the independent variable $1/C_i$ in Equation 82, data from stages with ammonia nitrogen concentrations less than 1 mg/l were not used. Table C-5 (Appendix C) contains the data used in the linear regression analyses. It was necessary to use the results obtained at 20°C during the first period of sampling (see Results and Discussion section) because, during the second period, the mixed liquor of most of the stages contained less than 1 mg/l of ammonia nitrogen. All of the ammonia removal data per stage are presented in Table C-6 (Appendix C). Table 41 contains a summary of the results obtained in the linear regression analyses for estimating the reaction rate and kinetic constant.

Table 41. Estimation of nitrification reaction rate and kinetic constant with Equation 82.

Mean[a] Temperature °C	n	Correlation Coefficient	K_N g/m²/d	K_N mg/l-N
15	5	0.945	2.439	0.76
19.8	8	0.966	4.624	4.68

[a]Calculated from the temperatures of stages which their data used in the linear regression.

n = number of determinations.

Figure 50 shows the measured concentrations of ammonia nitrogen in the RBC units operating at 15°C [Table C-6 (Appendix C)] plotted versus the kinetic constants obtained from the linear regression analyses. The lower part of the prediction curve does not pass through the measured data, indicating that there may be a threshold concentration below which ammonia nitrogen removal does not occur. Several explanations are possible for this phenomenon. For one, Caperon and Meyer (1972) postulated from experiments on the uptake of nitrogen compounds by phytoplankton that a minimum amount of intercellular nutrient is necessary to sustain growth. This minimum amount of intercellular nutrient was found to be related to the minimum concentration of nutrient in the bulk solution, S_o. Equation 83 was used to determine the maximum specific growth rate and half saturation constant under these conditions (Caperon and Meyer 1972).

$$V = \frac{V_M (S-S_o)}{K + (S-S_o)} \quad (83)$$

in which

V = nutrient uptake rate, 1/t

V_M = maximum nutrient uptake rate, 1/t

S = substrate concentration in solution, M/L^3

S_o = minimum substrate concentration, M/L^3

K = half-saturation constant, M/L^3

An other possible explanation is mass transport resistance which causes the mass flux of ammonia nitrogen from the solution into the attached growth to equal zero below a minimum ammonia nitrogen concentration in the solution.

Other explanations might account for this phenomenon. Insufficient analytical sensitivity is one

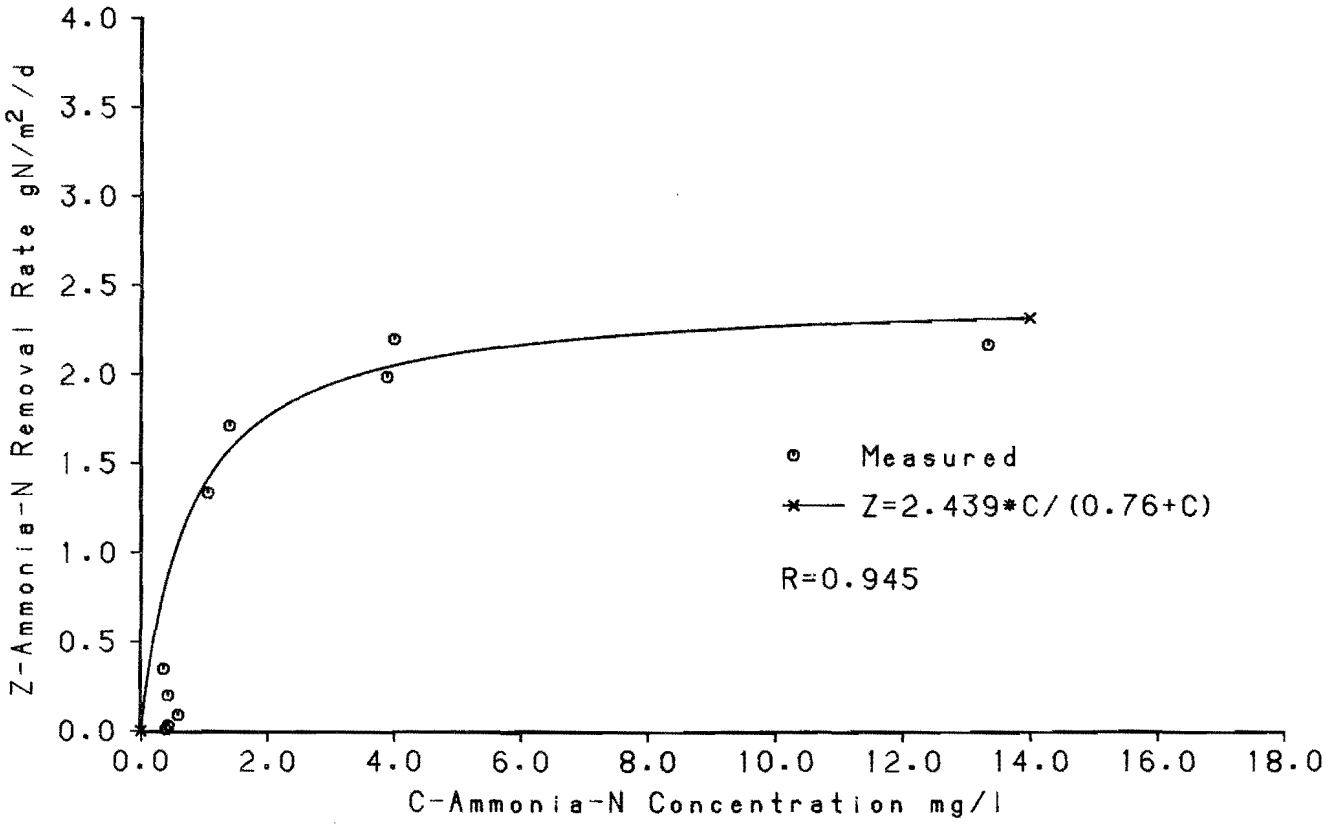


Figure 50. Relationship between the ammonia nitrogen removal rate and the ammonia nitrogen concentration at 15°C.

possibility. Nonbiodegradable ammonia nitrogen is another; however, this was not observed in the measurements conducted at 20°C. These possibilities are considered less likely than the two mentioned above.

The results at 15°C show the minimum concentration of ammonia nitrogen necessary to maintain the process to be approximately 0.4 mg/l-N. Equation 82 was modified to incorporate this concentration.

$$\left[\frac{Q(C_{i-1}-C_i)}{A_i}\right]^{-1} = \frac{K_N}{k_N}\frac{1}{(C_i-0.4)} + \frac{1}{k_N} \qquad (84)$$

A linear regression analysis with Equation 84 and the same data used to make the estimates reported in Table 41 gave a correlation coefficient = 0.970 (significance level = 1%); maximum reaction rate, k_N, g/m^2/d = 2.334; and the half saturation constant, K_N, mg/l = 0.45. Figure 51 shows the curve plotted using Equation 85, the curve for the unmodified relationship (Figure 50) and the measured values.

$$\frac{Q(C_{i-1}-C_i)}{A_i} = \frac{2.334\ (C_i-0.4)}{0.45 + (C_i-0.4)} \qquad (85)$$

The modified relationship of Equation 84 provides a better approximation of the measured data.

Figure 52 shows a plot of the data measured at 20°C and the curve plotted using the kinetic constants obtained from a linear regression analysis of Equation 82 (i.e., k_N = 4.624; K_N = 4.68). The plot of Equation 82 deviates from measured data points at the higher concentrations of ammonia nitrogen. Linear regression analyses of the Monod growth equation does not necessarily provide the best fit, although it is the best fit of the linearized form. One reason is that the low and medium concentrations have more impact than the high concentration on the determination of the intercept and the slope.

An attempt was made to improve the fit of the theoretical expression and the measured data by choosing the pair of kinetic constants which yield the minimum sum of squares (SSQ) between the predicted and observed values. Values of k_N in the range 3.00 to 4.60 gN/m^2/d and K_N values from 1.0 to 4.6 mg/l N were evaluated. The minimum SSQ was obtained using the values of k_N = 3.74 g/m^2/d and K_N = 2.8 mg/l. The SSQ was 1.062, and in comparison the previous set of constants obtained by linear regression yielded a SSQ of 1.502. Figure 53 shows the curve plotted using the resulting Equation 86 along with the previous equation and the measured data.

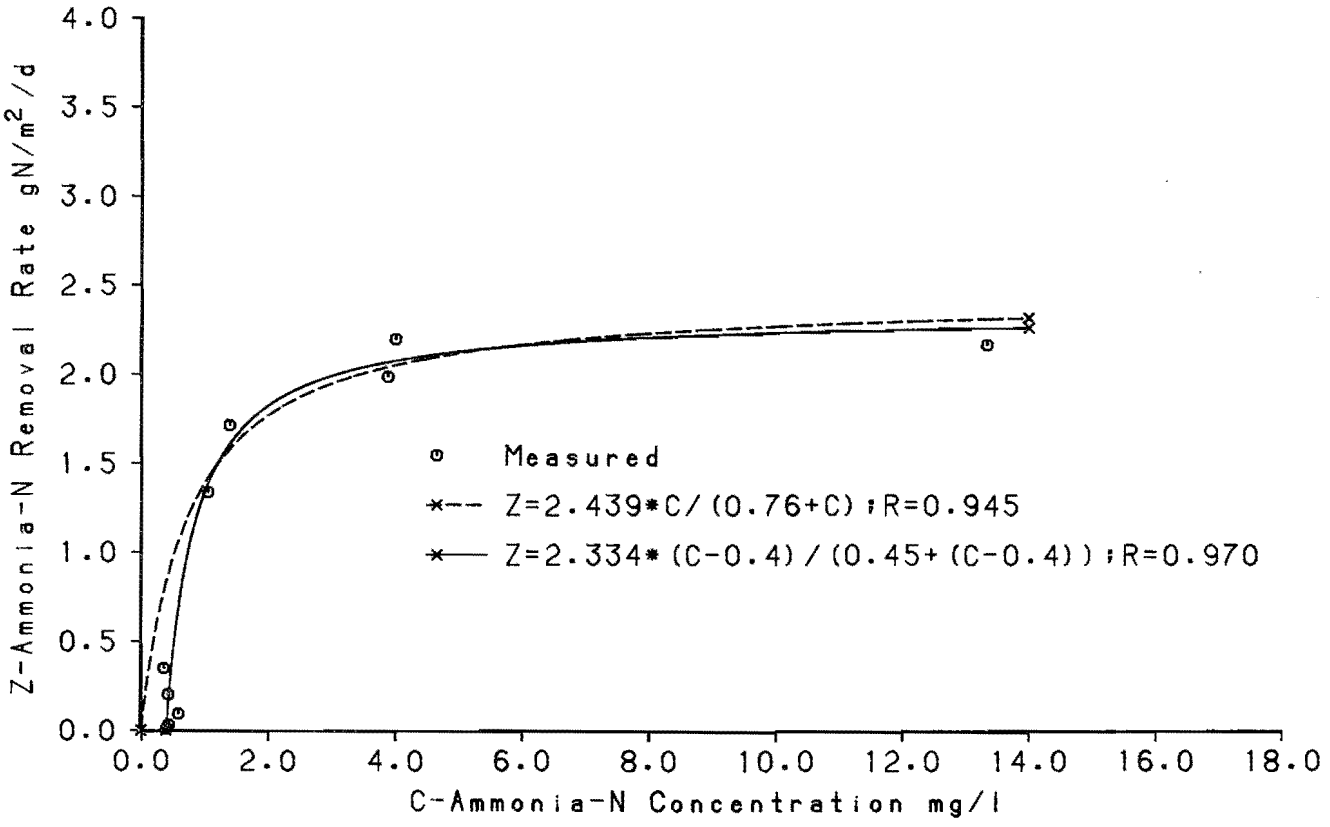


Figure 51. Comparison of the two predictive equations showing the relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 15°C.

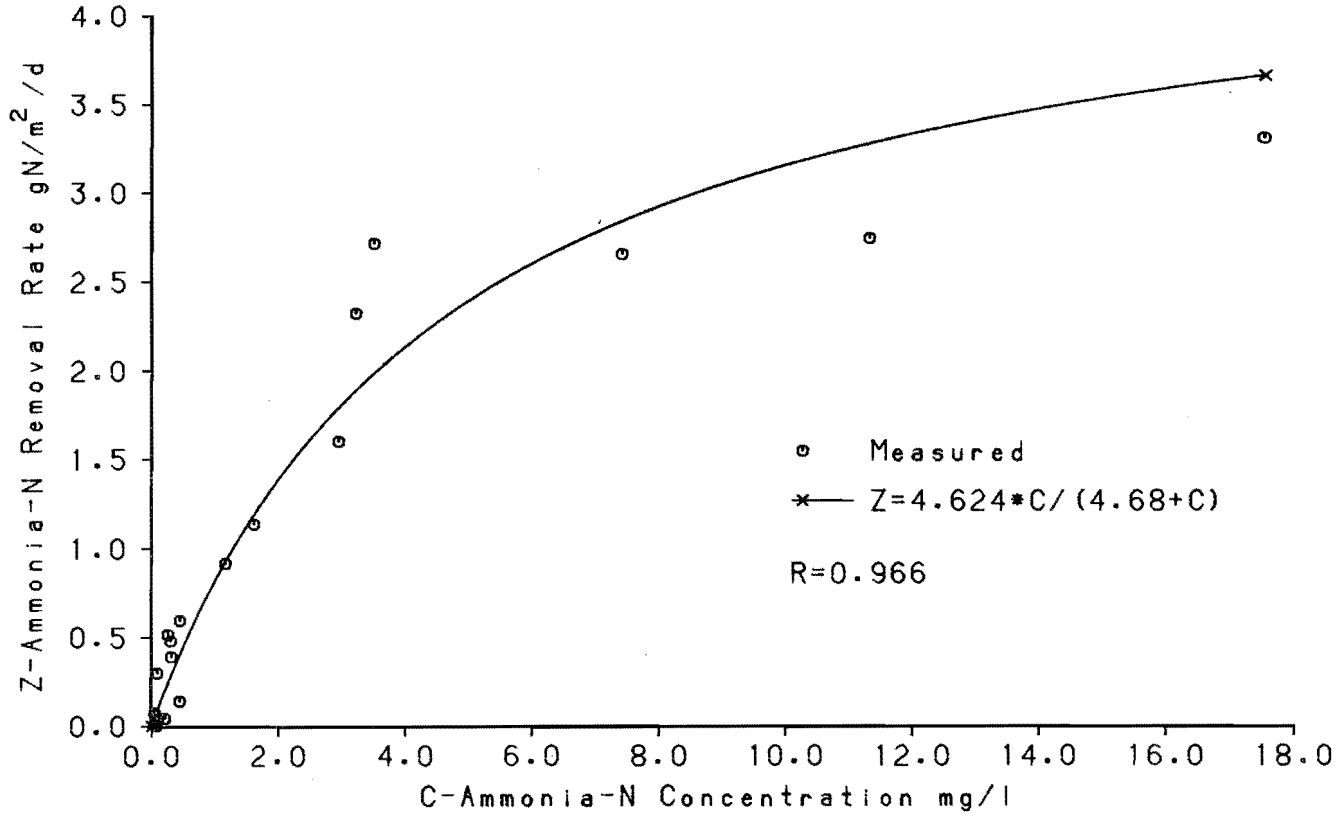


Figure 52. Relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 20°C.

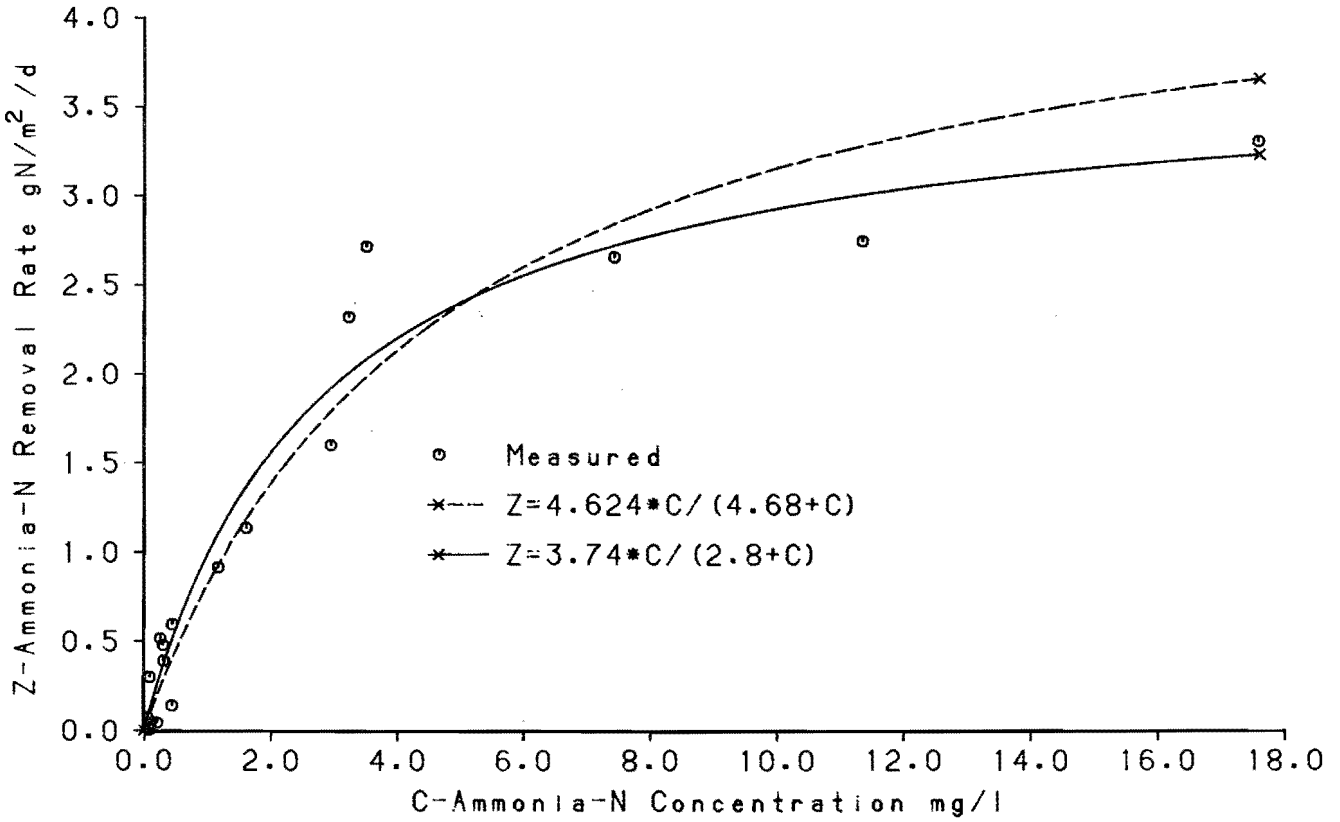


Figure 53. Comparison of the two predictive equations showing the relationship between ammonia nitrogen removal rate and the ammonia nitrogen concentration at 20°C.

$$\frac{Q(C_{i-1}-C_i)}{A_i} = \frac{3.74\ C_i}{2.8+C_i} \quad (86)$$

The data plotted in Figure 53 indicate that the predicted curves should pass through the origin. In other words, there was no indication that a minimum concentration of ammonia nitrogen was necessary at 20°C to maintain growth. A possible explanation is that at higher temperatures the mass transport resistance decreases, and as a result, the requirement for stored material is less. Table 42 contains a summary of the kinetic constants best representing the ammonia nitrogen removal data.

The relationship between the maximum ammonia nitrogen reaction rates and temperature can be expressed as follows:

$$(k_T) = (k_N)_{20}(\theta_N)^{T-20} \quad (87)$$

Using the data presented in Table 42 to obtain a fit of Equation 87 resulted in values of $(k_N)_{20} = 3.740$, and $\theta_N = 1.103$. The relationship between temperature and Nitrosomonas growth in river water reported by Parker et al. (1975) was equivalent to the relationship above. Parker et al. (1975) reported the function as $e^{0.098(T-15)}$. The temperature factor $e^{0.098}$ is equal to 1.103, which is equal to the θ_N value obtained in this study. Because of the base temperature of 15°C, the two expressions are not equivalent, but indicate excellent agreement for two such divergent systems.

Using the Arrhenius equation (Equation 56) to establish the relationship between temperature and the ammonia nitrogen reaction rates yielded a value for activation energy of 16,480 cal/mole. This value for activation energy is higher than the value of 13,900 cal/mole reported by Murphy et al. (1977); however, Murphy's value was based on TKN (total Kjeldahl nitrogen) rather than ammonia nitrogen. Mueller et al. (1980) chose a value of θ of 1.1 in the calibration of a theoretical model. The half-saturation constant of 0.45 mg/l obtained at 15°C is typical for nitrification (Parker et al. 1975; Painter 1970). With RBC systems, Saunders et al. (1980) determined K_N to be 0.18 to 1.0 mg/l. The half-saturation constant is also expected to increase with the temperature (Parker et al. 1975; Painter 1970).

Table 42. Summary of kinetic constants for ammonia nitrogen removal.

Temperature °C	Maximum Reaction Rate-gN/m²/d	Half Saturation Constant-mg/l-N	C_{min}[a] mg/l-N
15.2	2.334	0.45	0.4
20.0	3.740	2.80	0.0

[a]Minimum ammonia nitrogen concentration in solution, below which growth stops.

Overall ammonia nitrogen removal

The kinetic constants were developed in the previous section for stages where the growth of nitrifiers was not restricted. Generally, nitrifiers grew in the second through the fourth stages of the RBC units; however, limited removal was observed in the first stages when the organic loading rate was low. The ratio of observed ammonia nitrogen removal in the first stage to the simulated maximum removal rate using the kinetic equations can be expressed as a function of the organic loading rate for each temperature. Specifically,

$$\left[\frac{Q(C_0-C_1)}{A_1}\right]_{observed} = f_1\left[\frac{Q(C_0-C_{1m})}{A_1}\right]_{simulated} \quad (88)$$

$$f_1 = f(M) \quad (89)$$

where

C_0 = influent ammonia nitrogen concentration in the wastewater, mg/l-N.

C_1 = first stage effluent ammonia nitrogen concentration, mg/l-N

C_{1m} = first stage simulated ammonia nitrogen concentration, mg/l-N

Q = influent flow rate, m^3/d

A_1 = first stage surface area, m^2

f_1 = proportionality factor ≤ 1

M = organic load, $g/m^2/d$

Table C-7 (Appendix C) contains a summary of the calculated maximum ammonia nitrogen removal rates, observed rates, and the f_1 values. The f_1 values and the overall organic loading rates are presented in Table 43.

The relationship between the organic loading rates and the f_1 values is shown in Figure 54. From

Table 43. Summary of first stage ammonia nitrogen removal rate factors.

Temperature °C	Unit	Overall Organic Load gCOD/m²/d	f_1
15	A	3.984	0.994
15	B	7.496	0.018
15	C	9.875	0.380
15	D	13,916	0.040
20	A	6.915	0.762
20	B	9.734	0.570
20	C	12.513	0.000
20	D	13.971	0.080

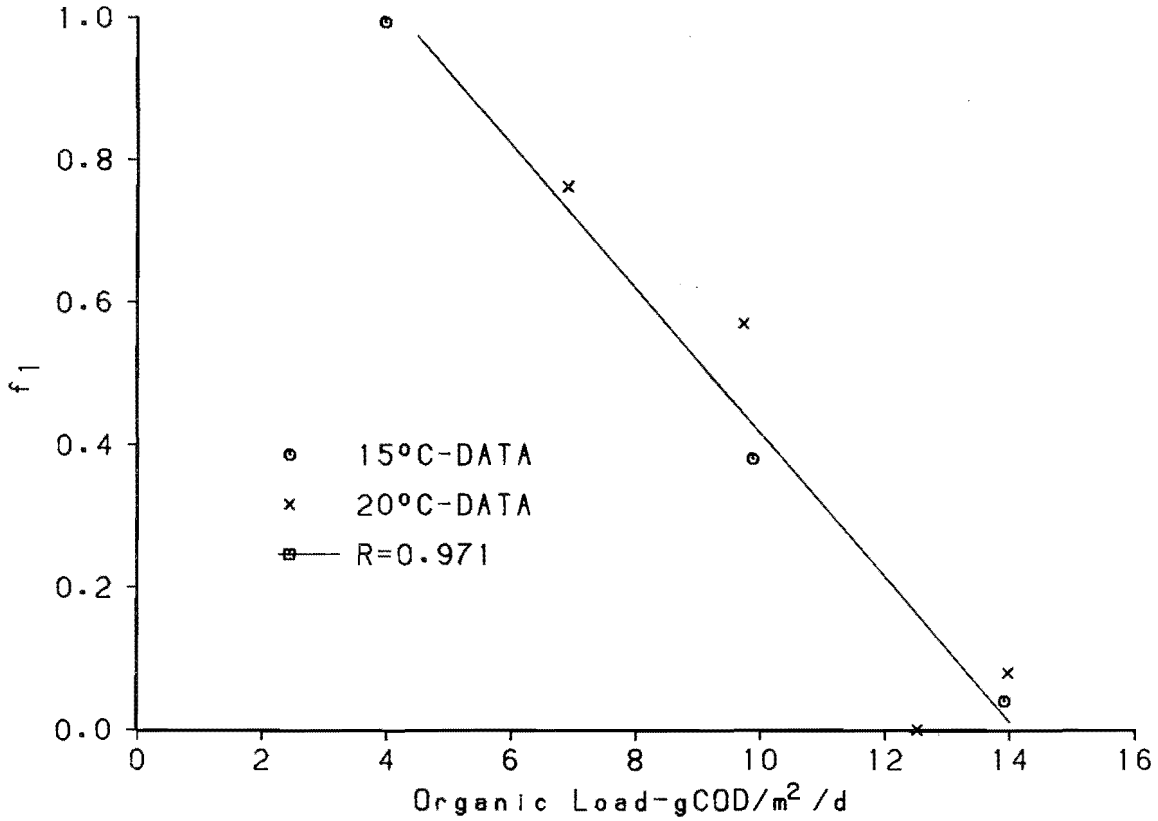


Figure 54. The relationship between the organic loading rate and the proportionality factor for ammonia nitrogen removal.

the location of the plotted points on Figure 54, one can see that f_1 values at 20°C were approximately 5-10 percent higher than the values obtained at 15°C. At both temperatures, the f_1 values decrease linearly as the organic loading rate increases. Excluding the value of f_1 and organic loading rate for unit B at 15°C, a linear regression analysis of the organic loading rate and the f_1 data at both temperatures results in a correlation coefficient of 0.971 (significance level = 1%). The linear relationship is:

$$f_1 = 1.43 - 0.1\ M;\ 4.3 \leq M \leq 14.3 \qquad (90)$$

Since f_1 is equal to or less than 1.0, the above limits are imposed on the organic loading rate. With Equation 90, it is possible to estimate the overall ammonia nitrogen removal rates as a function of the organic loading rate assuming that the growth of the nitrifiers is not restricted in the second stages. Specifically,

$$\frac{Q(C_0-C_4)}{\sum_{i=1}^{4} A_i} = k_N \left[\frac{f_1 C_{1m}}{K_N+C_{1m}} + \frac{C_2}{K_N+C_2} + \frac{C_3}{K_N+C_3} + \frac{C_4}{K_N+C_4} \right] \qquad (91)$$

Equation 91 can be solved for C_4 by determining C_{1m}, C_2 and C_3 stage by stage as shown in Table C-8 (Appendix C) where C_{1m} = ammonia nitrogen concentration at simulated conditions of maximum nitrification.

ENGINEERING SIGNIFICANCE

The steady-state kinetic models proposed and calibrated in this study provide rational quantitative relationships for design of an RBC process for treating domestic wastewater. The mathematical expressions provide a basis to calculate the RBC surface area required to meet prescribed effluent standards for carbonaceous substrate concentrations and ammonia nitrogen concentrations at temperatures ranging from 5°C to 20°C. The biomass yield and decay rate constants determined as a function of temperature enable one to estimate the mean suspended solids (SS) concentrations in the influent to the secondary clarifier.

Using the kinetic constants developed for 20°C in conjunction with Equations 55, 60, and 63 for given values of the hydraulic loading rate and the influent COD concentration, the effluent substrate concentration was calculated. Then, the removal efficiency was calculated using the influent concentration and the calculated effluent concentration. Design curves were then plotted for carbonaceous substrate removal in a four stage RBC process at 20°C as presented in Figure 55. The correction curves for adjusting for temperature are presented in Figure 56. The same calculations used to develop the curves for 20°C were performed with 5°C and 15°C kinetic constants. For equivalent removal efficiency at 20°C, 15°C, or 5°C, the ratios of the hydraulic loading rates at 20°C to those at 5°C or 15°C were used as the temperature factor. The mean temperature factor values for an influent COD of 200, 300, and 400 mg/l were plotted for a given efficiency versus the corresponding temperature. A design example to illustrate use of these curves is presented below.

A design flow rate of 3800 m^3/d (1 mgd) of domestic wastewater with a primary effluent COD concentration of 300 mg/l COD is to be treated with an RBC system to a degree that will produce a final effluent of 45 mg/l COD, or 85 percent removal. The design temperature is 10°C. Using Figure 55, the hydraulic load is found to be 1.75 gpd/sq ft (0.07

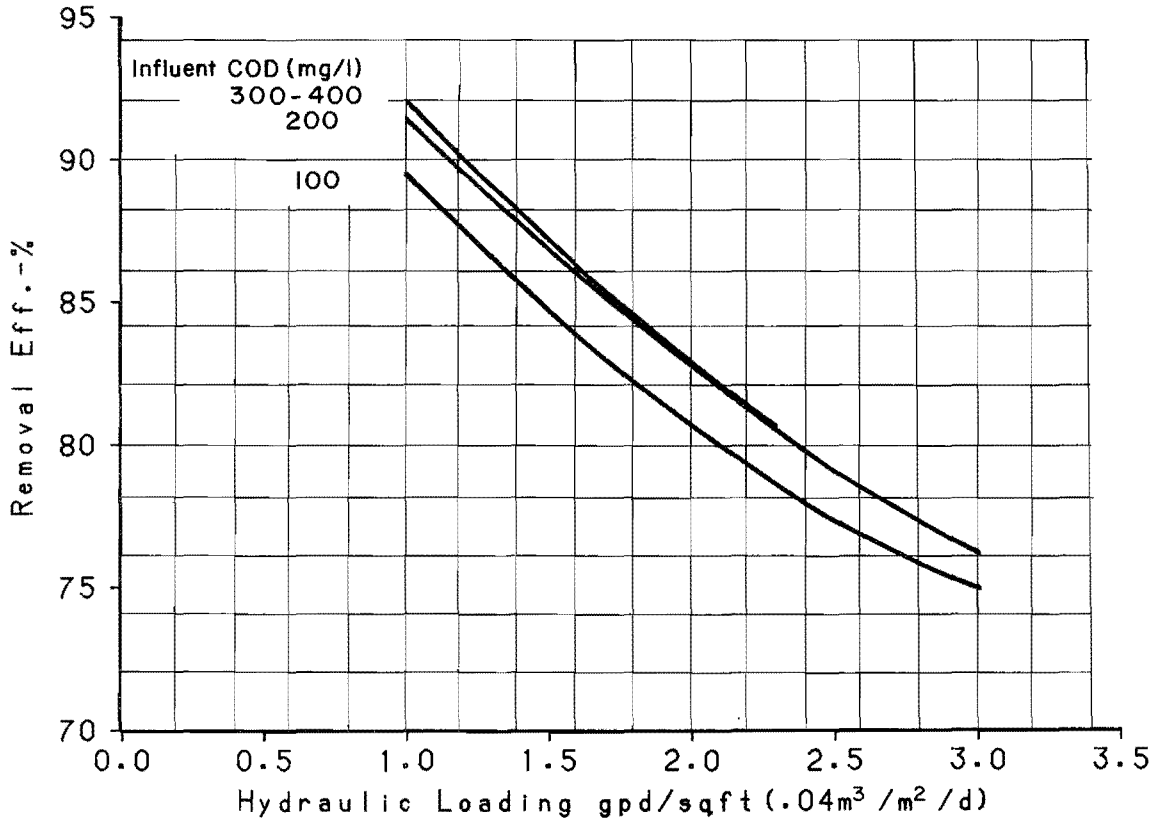


Figure 55. Design chart for COD removal in domestic wastewater treatment at 20°C.

$m^3/m^2/d$) at 20°C. At 10°C and 85 percent COD removal, the temperature factor is 2.4 (Figure 56). To meet the required effluent quality at 10°C, the designed hydraulic loading rate cannot exceed 1.75/2.4 or 0.03 $m^3/m^2/d$ (0.73 gpd/sq ft). The required total effective contactor area will be:

$$3800/0.03 = 127{,}000\ m^2\ (1.365 \cdot 10^6\ ft^2)$$

For a constant influent COD concentration, various hydraulic loading rates and influent ammonia concentrations were assumed, and the effluent ammonia concentrations were calculated using Equations 90 and 91. The kinetic constants were obtained from Equation 87. The half saturation constant, K_N, and C_{min} were taken from the experiments at 5°C, 15°C, and 20°C. Three design curves based on the above equations are presented in Figures 57, 58, and 59 for an influent COD concentration of 300 mg/l and for temperatures of 10°C, 15°C, and 20°C, respectively. For other influent COD concentrations, similar design curves can be developed using the equations presented in this study. Figure 59 shows that for a hydraulic loading rate of 0.07 $m^3/m^2/d$ (1.75 gpd/sq ft) and an influent ammonia nitrogen concentration of 20 mg/l, the expected ammonia nitrogen removal at 20°C will be 94 percent leaving 0.6 mg/l ammonia nitrogen in the effluent. Figure 58 shows that at a hydraulic loading rate of 0.03 $m^3/m^2/d$ (0.73 gpd/sq ft) and 10°C, the expected ammonia nitrogen removal will be about 97.5 percent, leaving 0.5 mg/l ammonia nitrogen in the effluent.

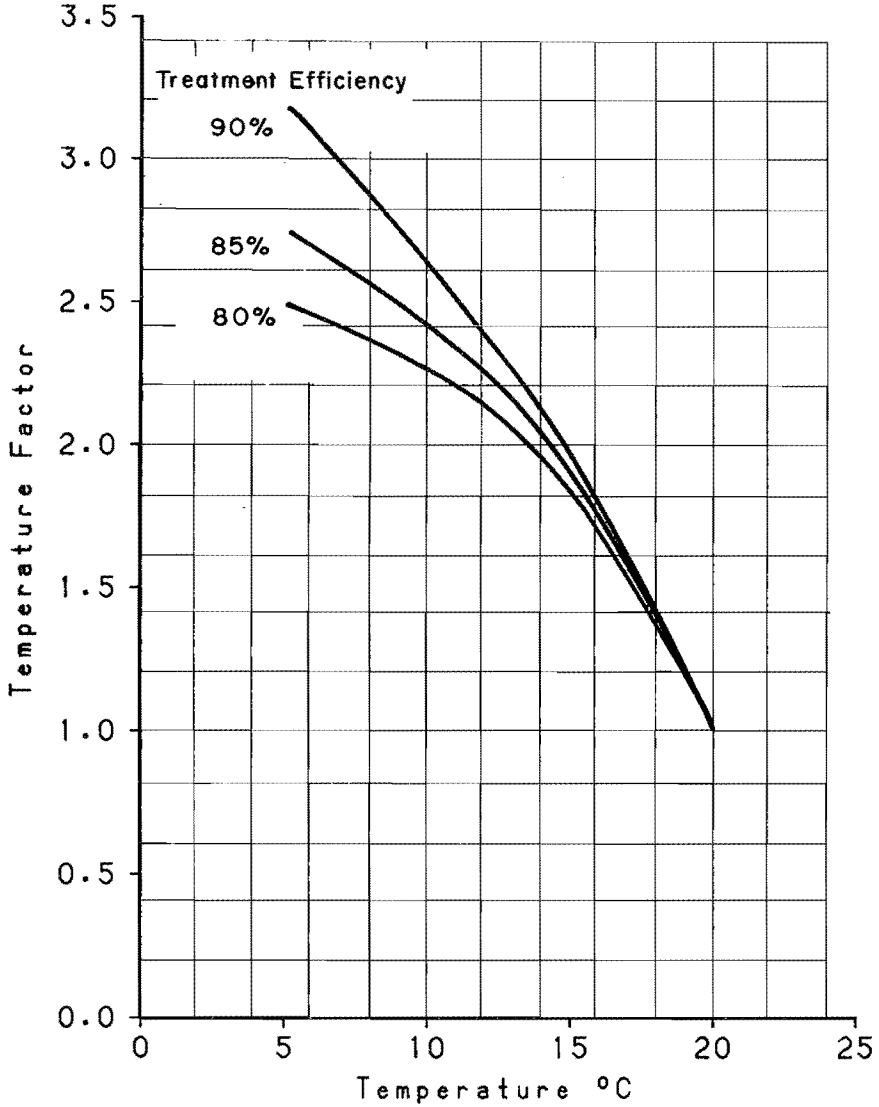


Figure 56. Temperature correction for COD removal.

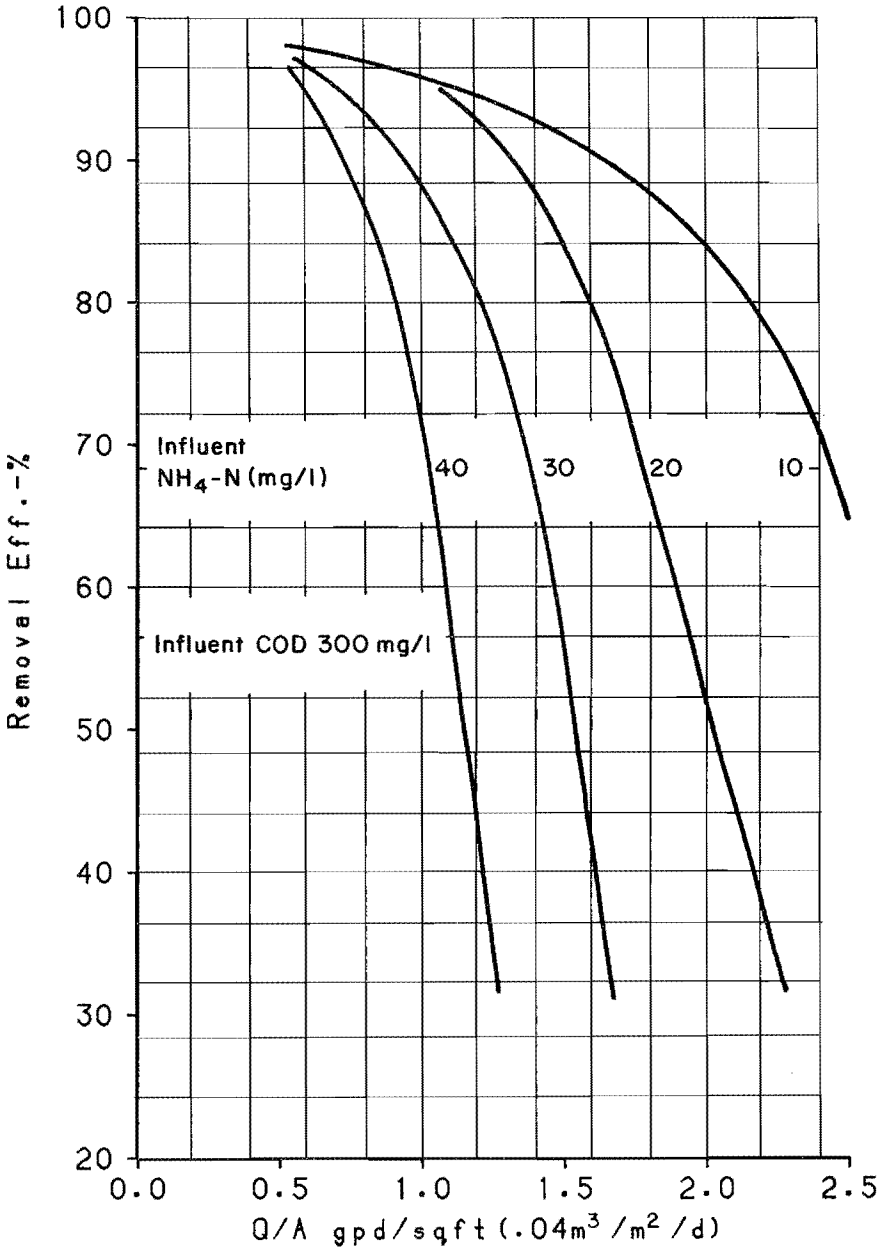


Figure 57. Design chart for ammonia nitrogen removal at 10°C.

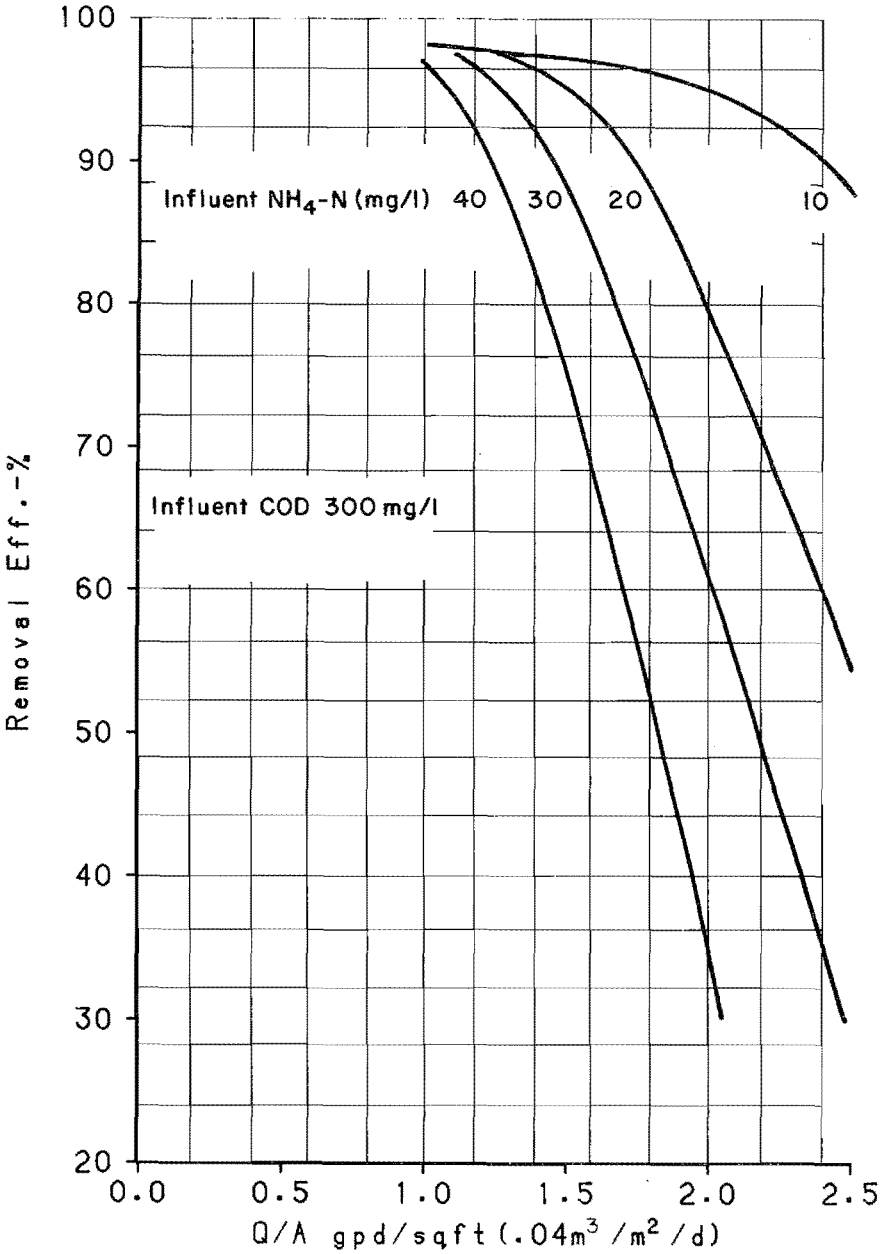


Figure 58. Design chart for ammonia nitrogen removal at 15°C.

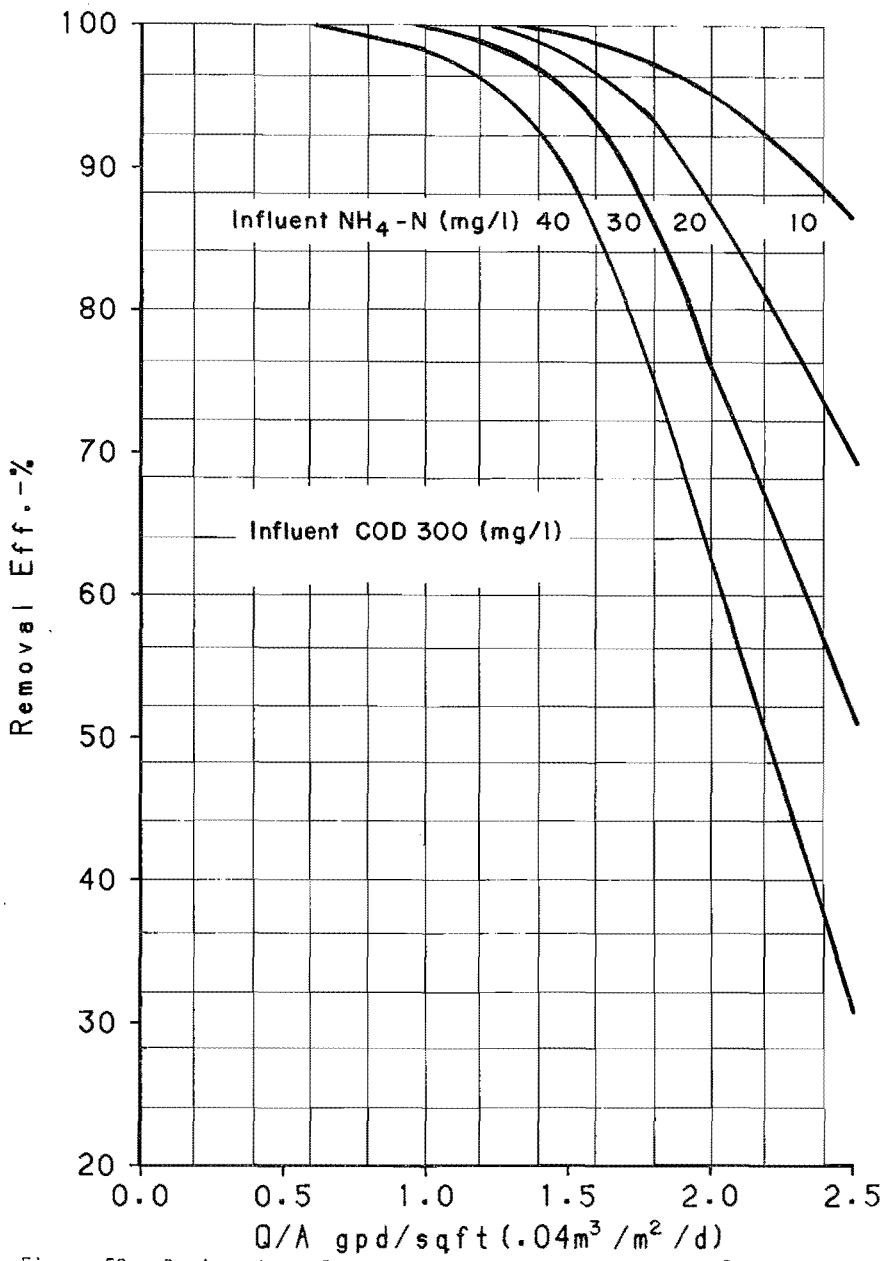


Figure 59. Design chart for ammonia nitrogen removal at 20°C.

SUMMARY, CONCLUSIONS, AND RECOMMENDATIONS

Four, four-stage, 38 cm diameter, laboratory scale rotating biological contactor (RBC) units were employed to gather empirical data for use in developing kinetic models of the RBC process for treating domestic wastewater. The wastewater used was a settled domestic wastewater with mean concentrations of chemical oxygen demand (COD) of 275 mg/l and ammonia nitrogen of 25 mg/l. The study was conducted at temperatures of 5°C, 15°C, and 20°C. At each temperature, the experimental units received constant hydraulic loading rates, and the hydraulic residence time in each unit was approximately 2 hours. At each temperature the experimental units were operated at four different organic loading rates, ranging from 4 $gCOD/m^2/d$ to 14 $gCOD/m^2/d$. The various organic loading rates were achieved by diluting the raw wastewater with dechlorinated tap water.

At each temperature, the experimental units achieved steady-state conditions after 4 weeks of operation. Operation after achieving steady-state conditions was continued for 2 to 6 weeks, to provide an adequate number of samples for process analysis and estimation of kinetic model parameters.

The experimental units were operated in an environment that assured adequate dissolved oxygen (DO) concentrations and pH values. The DO concentrations in the mixed liquor ranged from 1.9 mg/l to 9.9 mg/l. The higher concentrations of DO were measured at the coldest temperature, and in the last stages of the RBC units. The pH values generally were in a range of 7.8 to 8.1.

In the first stages of the RBC units, a dense and spongy attached biomass developed, and the attached biomass was stable during the steady-state period. In the succeeding stages, the amount of attached growth was less and had a smoother appearance. In the last stages the growth was patchy and unstable. The amount of attached biomass was dependent upon temperature and influent organic loading rate. A saturation type function described the relationship between the quantity of attached biomass and the influent organic loading rate.

The percent COD removal was influenced by the temperature and the organic loading rate. The overall substrate (influent total COD minus effluent filtered COD) removal efficiencies were 80 percent, 85 percent, and 90 percent for 5°C, 15°C, and 20°C, respectively. The majority of the COD removal occurred in the first stages, and the first stage removals were 77 percent, 80 percent, and 85 percent for 5°C, 15°C, and 20°C, respectively. Sludge production in the fourth stage effluent was 0.557 grams suspended solids (SS) per gram COD removed at 5°C. At 15°C and 20°C, the sludge production was 0.430 grams SS per gram COD removed.

Different processes were occurring in the first stages than in the following stages. In the first stage, the major process was carbonaceous substrate removal, and in the other stages, stabilization of sloughed biomass and nitrification were observed. Because of the instability of the attached growth (inconsistent sloughing) in the last stages, in some cases it was difficult to show that biomass stabilization was occurring.

In the experiments conducted at 5°C, significant ammonia nitrogen conversion or removal was not observed. At higher temperatures significant ammonia removal was observed beyond the second stages. The removal rates generally corresponded with nitrification rates and were related to mixed liquor temperature and ammonia nitrogen concentrations. The maximum ammonia nitrogen removal rates were about 2 $grams/m^2/d$ and 2.7 $grams/m^2/d$ at 15°C and 20°C, respectively. Overall ammonia nitrogen removal ranged from 87 percent to 98 percent at 15°C, and from 91 percent to 99 percent at 20°C. As the influent organic loading rates increased, the overall ammonia nitrogen removal decreased.

The volatile fraction of the attached growth ranged from 60 percent to 70 percent. Generally, the lowest volatile fraction of the attached biomass was observed in the last stages. The total organic carbon in the attached growth ranged from 30 percent to 42 percent, total nitrogen ranged from 6 percent to 8 percent, and total phosphorus ranged from 1.4 to 1.7 percent.

A kinetic model was developed for each major process of the RBC system, and the experimental data were used to calibrate the models. The temperature dependence of the kinetic constants was formulated using typical expressions employed in other wastewater processes [$k = k_{20}\ \theta^{(T-20)}$].

The yield coefficients expressed as grams VS produced per gram COD removed were found to be 0.66, 0.80, and 0.63 for 5°C, 15°C, and 20°C, respectively. A decay coefficient of 0.27/d at 20°C and a temperature factor, θ_d, of 1.09 were obtained. The reaction rates for the stabilization of sloughed biomass were approximately the same as the decay rates.

The first stage carbonaceous substrate removal was described using Monod growth kinetics. The half-saturation constants ranged from 61.6 to 276.4 mg/l COD, with the lowest value obtained for the 5°C experiments. The maximum specific growth rate was 1.87/d at 5°C and approximately 6/d at the higher temperatures. The maximum reaction rate was 9.50/d at 20°C and the temperature factor, θ_s, was 1.09. The quantity of attached growth in the first stages could be predicted using a saturation type function, and the quantity of attached biomass increased as the temperature increased. Because of the low removal of carbonaceous substrate in the second through fourth stages rational models could not be used to describe

the latter stages of the process. A multiple regression analysis was used to relate temperature, substrate concentration, and reaction rates in the second through the fourth stages. The reaction order in the latter stages was estimated to be 0.763, and the temperature factor, θ_L, was 1.11.

The ammonia nitrogen removal kinetics were also described using Monod growth kinetics. The half-saturation constants were 0.45 mg/l-N and 2.8 mg/l-N at 15°C and 20°C, respectively. The maximum reaction rate at 20°C was calculated to be 3.740 $gN/m^2/d$ with a temperature factor, θ_N, of 1.10. The inhibition of ammonia nitrogen removal in the first stages was correlated with the influent organic loading rate.

Based upon the results of this study, the following major conclusions can be derived for a four-stage RBC process treating domestic wastewater:

1. The RBC process can achieve 90 percent removal of carbonaceous substrate and ammonia nitrogen at 20°C receiving an influent organic loading rate of 14 $gCOD/m^2/d$.

2. The removal efficiency of both COD and ammonia nitrogen decreased when the temperature was below 20°C.

3. The suspended solids production was 0.43 grams per gram of COD removed at 15°C and 20°C and production increased to 0.557 at 5°C.

4. The yield coefficients and specific maximum growth rates indicated that the optimum growth conditions for attached biomass occur around 15°C.

5. The chemical composition of the attached growth in the RBC units was similar to other types of bacteria, and the ratio of C:N:P in the attached growth was 28:5.2:1.

6. The quantity of attached growth was predicted by a saturation type relationship incorporating a temperature factor of 1.015.

7. Decay rates for the attached biomass were affected by the temperature and a temperature factor, θ_d, of 1.09 was observed.

8. Carbonaceous substrate removal kinetics in the first stage were described with Monod growth kinetics by incorporating the total attached biomass.

9. The maximum reaction rates for carbonaceous substrate removal were related to temperature, and a temperature factor, θ, of 1.09 was observed in the first stages and a value of 1.10 in the succeeding stages.

10. The increase of the half-saturation constant (K_s) for carbonaceous substrate removal from 5°C to 15°C was greater than the increase from 15°C to 20°C.

11. Carbonaceous substrate removal in the second through fourth stages receiving an effluent from the first stage containing a filtered COD concentration of approximately 50 mg/l was described with an equation having a reaction order of 0.763.

12. Estimated reaction rates for the removal of particulate COD in the second through fourth stages indicate that the process was carried out by endogenous respiration of the stored substrate in the sloughed biomass via reattachment in the succeeding stages.

13. Ammonia nitrogen removal was achieved principally through nitrification.

14. The kinetics of ammonia nitrogen removal were described with Monod growth kinetics.

15. The temperature factor, θ_N, for the relationship between the maximum reaction rate for ammonia nitrogen removal and temperature was found to be 1.10.

16. Below 20°C, ammonia nitrogen removal did not readily occur when the minimum ammonia nitrogen concentration was less than 0.4 mg/l.

17. The half-saturation constants for ammonia nitrogen removal were 0.45 mg/l and 2.8 mg/l for temperatures of 15°C and 20°C, respectively.

18. At a temperature of 5°C there was no significant ammonia nitrogen removal.

19. Ammonia nitrogen removal rates in the first stages were inversely proportional to the influent organic loading rate.

20. The kinetic constants developed in this study provide a rational design approach for RBC systems and are summarized in Table 44.

RECOMMENDATIONS

In order to increase the usefulness of this study, it is recommended that the following additional work be completed:

1. Conduct experiments to determine the kinetic constants at 10°C and at 25°C.

2. Conduct experiments with higher influent flow rates and/or higher substrate concentrations than used in this study in order to develop rational models in the last stages of the RBC process.

3. Study the dynamics of sloughing and sloughed biomass stabilization.

4. Determine the kinetic constants based upon the removal of biochemical oxygen demand (BOD_5).

5. Perform dissolved oxygen balances in the RBC system as a function of influent organic loading rate, detention time, and temperature.

Table 44. Summary of the kinetic constants.

Process	Parameter	Temperature(°C) or range	Recommended value or range
Biomass yield	Y = Yield coefficient (gVSS/gCOD)	15 5, 20	0.80 0.63 - 0.66
Biomass decay	k_d = Decay coefficient (1/d)	20	0.27
	θ_d = Temperature factor	5 - 20	1.09
Carbonaceous substrate removal kinetics - 1st stage	k = Maximum reaction rate (1/d)	20	9.50
	θ_s = Temperature factor	5 - 20	1.09
	K_s = Half saturation constant (mg/ℓ COD)	5 15 - 20	61.6 262 - 276
	k_x = Maximum attached growth gVS/m^2	20	56.9
	θ_x = Temperature factor	5 - 20	1.015
	K_x = Constant	5 - 20	23.77 - 31.07
Carbonaceous substrate removal kinetics - last stages	k_L = Maximum reaction rate $(g/m^2/d)$	20	0.0444
	θ_L = Temperature factor	5 - 20	1.11
	n = Apparent reaction order	5 - 20	0.763
Ammonia-N removal - last stages	k_N = Maximum reaction rate $(g/m^2/d)$	20	3.74
	θ_N = Temperature factor	15 - 20	1.1
	K_N = Half saturation constant (mg/ℓ - NH_4-N)	15 - 20	0.45 - 2.8
	c_m = Minimum ammonia-N concentration required (mg/ℓ - NH_4-N)	15 - 20	0 - 0.4
Ammonia-N removal - first stage (and second stage in high organic loadings)	f_1 = Fraction of the ammonia-N removal rate at unrestricted nitrifier growth conditions	15 - 20	0 - 1.0
	Organic load dependent		-0.1

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APPENDIX A

HYDRAULIC CHARACTERISTICS OF THE EXPERIMENTAL RBC UNITS

The hydraulic characteristics of a single stage of the experimental RBC units were determined using sodium chloride (NaCl) as a tracer. The experiment was designed to simulate the operating conditions employed in the kinetic studies. The volume of the stage was maintained at 6.1 liters and tap water was added to the system at flow rates of 200 to 205 ml/min. The mean theoretical detention time (t_o) at a volume of 6.1 liters was 30.1 minutes. Four clean discs were mounted on the shaft and rotated at a speed of 16 rpm. The tap water was pumped from a 1.9 m^3 (500 gal) storage tank. An initial mass of 30.3 grams of NaCl was injected at the inlet to the stage, and the initial concentration (C_o) of NaCl was 4.967 g/l. The initial electrical conductivity at 25°C (EC_{25}) in the stage was 8950 µmho. The EC_{25} of tap water was 325 mho The calibration curve for the relationship between salt concentration and the electrical conductivity (EC) was prepared using the same source of tap water for dilution. The response of the system to the impulse disturbance was measured in terms of electrical conductivity. All electrical conductivity measurements were corrected to 25°C. Figure A-1 shows the response of the system to the injection of NaCl and the theoretical curve for the response of a complete mix reactor to a single injection of a tracer. An examination of Figure A-1 shows that the hydraulic characteristics of the stage are represented by a complete-mix flow model. The mathematical expression used to describe the response to the injection of a slug of tracer in a complete mix flow reactor is as follows:

$$C/C_0 = \exp(-t/t_o)$$

in which

C = salt concentration at time t, M/L^3

C_o = initial salt concentration, M/L^3

t = time, minutes

t_o = theoretical detention time, minutes

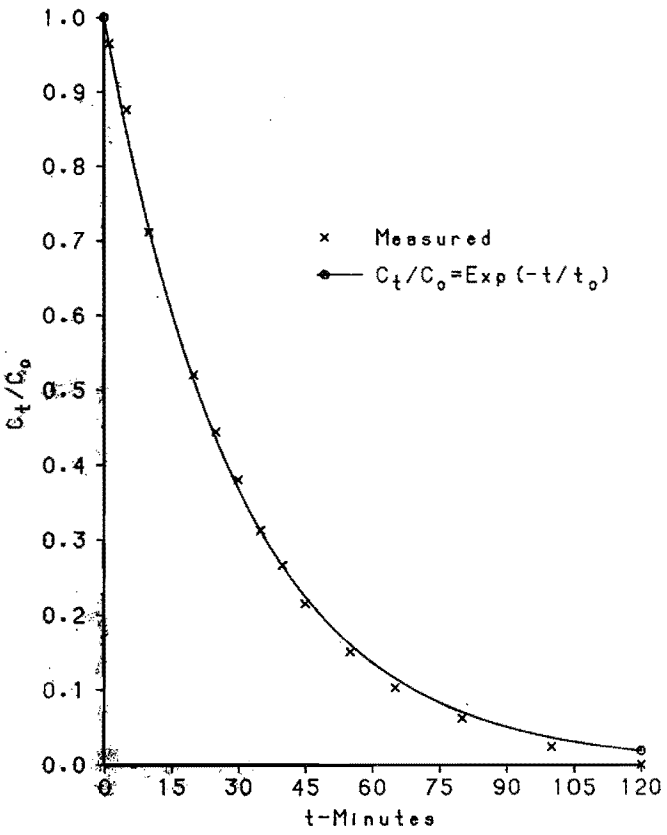

Figure A-1. Theoretical concentration-time curve for the injection of a slug of tracer into a complete mix reactor and the measured response of the injection of a tracer into a single stage of the experimental RBC units.

APPENDIX B

EXPERIMENTAL DATA

Table B-1. Influent flow rates and the percentage of wastewater applied to the four experimental RBC units when operating at a temperature of 5°C.

Unit	A		B		C		D
Date	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min
5/28	200	67.5	200	50.0	205	80.5	200
5/29	195	66.7	195	51.3	205	80.5	195
6/4	205	65.9	200	50.0	205	80.5	210
6/6	200	65.0	200	50.0	210	81.0	205
6/18	200	67.5	195	48.7	210	78.6	210
6/20	200	65.0	190	50.0	210	78.6	210
6/23	200	65.0	200	47.5	205	78.1	215
6/24	200	65.0	200	47.5	205	78.1	215
6/25	205	63.4	200	47.5	205	78.1	210
6/26	200	65.0	200	47.5	205	78.1	215
6/30	200	65.0	195	48.7	200	80.0	215
7/1	195	66.7	200	47.5	200	80.0	215
7/2	200	65.0	200	45.0	200	80.0	215
7/3	200	67.5	200	47.5	200	80.0	215
7/4	195	66.7	195	48.7	205	78.1	210
7/5	195	69.2	195	48.7	205	80.5	210
7/6	195	69.2	195	48.7	200	80.0	210
7/7	200	67.5	195	48.7	200	80.0	210
7/9	195	69.2	195	48.7	205	80.5	215

*Unit D received undiluted wastewater, i.e., 100 percent sewage.

Table B-2. Influent flow rates and the percentage of wastewater applied to the four experimental RBC units when operating at a temperature of 15°C.

Unit	A		B		C		D
Date	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min
12/10	200	25.0	200	50.0	195	71.8	200
12/11	190	26.3	195	48.7	195	71.8	195
12/12	185	27.0	200	47.5	190	71.1	200
12/13	200	25.0	195	48.7	200	75.0	200
12/14	190	26.3	200	50.0	195	71.8	195
12/15	190	26.3	190	50.0	185	73.0	195
12/18	190	26.3	200	50.0	195	74.4	200
12/19	190	26.3	200	50.0	200	75.0	200
12/20	200	25.0	200	50.0	200	75.0	200
12/21	200	25.0	200	50.0	200	75.0	200
12/24	-	-	-	-	195	71.8	210
12/25	-		-		200	72.5	210

*Unit D received undiluted wastewater, i.e., 100 percent sewage.

Table B-3. Influent flow rates and the percentage of wastewater applied to the four experimental RBC units when operating at a temperature of 20°C.

Unit	A		B		C		D
Date	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min	% Wastewater	Flow Rate mℓ/min
3/19	195	48.7	200	67.5	195	87.2	200
3/20	205	48.8	195	71.8	205	87.8	195
3/26	190	50.0	200	70.0	195	84.6	200
3/27	195	48.7	200	70.0	195	84.6	200
3/31	200	50.0	195	69.2	200	85.0	205
4/2	195	48.7	195	69.2	205	82.9	205
4/3	190	50.0	195	69.2	200	87.5	205
4/5	195	48.7	195	71.8	200	87.5	205
4/7	195	48.7	200	67.5	200	87.5	200
4/8	195	48.7	200	67.5	195	87.2	205
4/9	195	48.7	195	69.2	210	83.3	205
4/10	195	48.7	195	69.2	200	87.5	205
4/11	195	48.7	200	67.5	195	87.2	205
4/12	195	48.7	195	69.2	200	87.5	205

*Unit D received undiluted wastewater, i.e., 100 percent sewage.

Table B-4. Mean liquid temperatures (°C) in the RBC units when the constant temperature room was maintained at 5°C.

Unit	First Stage			Second Stage			Third Stage			Fourth Stage		
(Period of Operation)	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.
B (5/26-7/9)	20	6.1	0.6	20	5.3	0.7	20	4.8	0.8	20	4.5	0.8
A (5/29-7/7	18	6.0	0.6	18	5.3	0.7	18	4.4	0.8	18	3.8	0.8
C (5/29-7/7)	18	5.9	0.6	18	5.1	0.7	18	4.6	0.8	18	4.2	0.8
D (5/28-7/6)	18	5.4	0.6	18	4.7	0.6	18	4.2	0.7	18	3.8	0.8

n = number of determinations

S.D. = standard deviation

Table B-5. Mean liquid temperatures (°C) in the RBC units when the constant temperature room was maintained at 15°C.

Unit (Period of Operation)	First Stage			Second Stage			Third Stage			Fourth Stage		
	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.
A (12/10-12/19)	8	16.7	0.4	8	15.6	0.5	8	15.0	0.4	8	14.8	0.4
B (12/12-12/21)	7	16.0	0.3	7	15.4	0.4	7	14.4	0.5	7	14.1	0.5
C (12/11-12/25)	11	16.0	0.4	11	15.0	0.4	11	14.4	0.5	11	14.1	0.4
D (12/10-12/25)	12	16.3	0.3	12	15.6	0.2	12	15.1	0.2	12	14.7	0.3

n = number of determinations

S.D. = standard deviation

Table B-6. Mean liquid temperatures (°C) in the RBC units when the constant temperature room was maintained at 20°C.

Unit (Period of Operation)	First Stage			Second Stage			Third Stage			Fourth Stage		
	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.	n	Average	S.D.
A (3/26-4/12)	7	20.7	0.5	7	20.0	0.4	7	19.1	0.4	7	18.6	0.6
B (3/27-4/11)	6	21.2	0.3	6	20.7	0.3	6	20.1	0.2	6	19.8	0.3
C (3/26-4/12)	7	20.8	0.3	7	20.3	0.3	7	20.0	0.4	7	19.3	0.3
D (3/20-4/11)	8	20.5	0.0	8	20.1	0.2	8	19.6	0.2	8	19.5	0.3

n = number of determinations

S.D. = standard deviation

Table B-7. Wastewater characteristics during period of operation at 5°C.

Date	COD mg/ℓ		SS mg/ℓ	VSS mg/ℓ	TKN mg/ℓ	NH_4-N mg/ℓ	NO_2-N mg/ℓ	NO_3-N mg/ℓ
	Total	Filtered						
5/28	121.5	95.7	64	64				
5/29	127.0							
6/4	205.4							
6/6	151.1	43.5	64	64				
6/18	170.0	55.1	88	78				
6/20	183.0	44.0	88	82				
6/23	255.0	89.0	90	90				
6/24	197.1		112	106				
6/25	135.0	59.3	58	58	50.5	20.16	0.155	1.40
6/26	106.0		42	42	53.3	21.05	0.166	1.44
6/30	205.9	48.1	76	76	32.0	17.79	0.240	<0.04
7/1	213.7	44.1	110	96	37.7	18.07	0.025	0.03
7/2	151.3		72	72				
7/3	219.0	36.7	70	70				
7/4	174.6	35.3	60	60				
7/5	165.3	54.8	64	64				
7/6	165.9	30.2	74	74	45.6	24.28	0.180	0.92
7/7	213.6	40.6	98	98	45.5	19.45	0.165	0.64
7/9	195.5	43.5	96	96				

Table B-8. Wastewater characteristics during period of operation at 15°C.

Date	COD mg/ℓ		SS mg/ℓ	VSS mg/ℓ	TKN mg/ℓ	NH_4-N mg/ℓ	NO_2-N mg/ℓ	NO_3-N mg/ℓ
	Total	Filtered						
12/10	286.0							
12/11	231.5	93.0	86	86			0.012	<0.04
12/12	388.0	156.0	138	110				
12/13	362.5	103.0	150	132	38	30.74	0.010	<0.04
12/14	419.0						0.013	0.04
12/15	346.0							
12/18	129.0	35.0	90	86		29.72	0.008	<0.04
12/19	273.0	98.0	90	90	43	28.92	0.005	<0.04
12/20	193.0	113.0	65	65				
12/21	230.0	82.0	82	72				
12/24	176.0	57.0	40	40				
12/25	148.0	47.0	86	86				

Table B-9. Wastewater characteristics during period of operation at 20°C.

Date	COD mg/ℓ		SS mg/ℓ	VSS mg/ℓ	TOC mg/ℓ		TKN mg/ℓ	NH_4-N mg/ℓ	NO_2-N mg/ℓ	NO_3-N mg/ℓ
	Total	Filtered			Total	Filtered				
3/19	167.0	101.0	120	112	56.8	34.0	54.80	34.45	0.170	0.13
3/20	195.0	56.3	46	38	49.4	24.1	48.45	36.90	0.140	0.41
3/26	248.6		80	80			42.20	32.02	0.385	0.42
3/27	211.4						57.00	36.46	0.027	0.07
3/31	264.8									
4/2	189.1				62.8	31.0	29.75	17.15	0.390	0.66
4/3	257.4				66.8		38.10	18.34	0.033	<0.04
4/5	296.5									
4/7	326.7	109.0	148	128	83.2		38.10	15.71	0.315	0.44
4/8	485.8				97.6		49.55	11.76	0.025	0.68
4/9	174.9	59.5	74	74			39.50	15.86	0.600	0.80
4/10	307.0						48.75	16.31	0.225	0.12
4/11	426.0	108.0	72	72						
4/12	377.0									

Table B-10. Mean characteristics of the influent to each RBC unit when operating at 5°C.

Unit	A			B			C			D		
Parameter	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
Flow rate ℓ/d	17	286.7	4.6	19	284.2	4.4	17	294.0	5.2	17	302.8	8.2
Total COD mg/ℓ	17	118.4	25.6	19	85.6	18.4	17	142.0	30.4	17	173.3	39.6
Filtered COD mg/ℓ	12	34.5	12.3	14	25.0	9.4	12	38.4	11.9	12	53.0	20.3
SS - mg/ℓ	15	51.5	13.1	17	37.7	9.3	15	61.6	15.6	15	75.5	19.2
VSS - mg/ℓ	15	49.1	13.5	17	35.2	6.5	15	60.3	16.5	15	71.5	10.6
TKN - mg/ℓ	6	29.2	5.2	6	21.2	3.7	6	35.0	6.0	5	43.8	8.9
Ammonia-N mg/ℓ	6	13.34	1.86	6	9.69	1.2	6	15.98	1.88	5	20.27	2.6
Nitrite-N mg/ℓ	6	0.103	0.047	6	0.075	0.035	6	0.123	0.057	5	0.153	0.079
Nitrate-N mg/ℓ	5	0.58	0.38	5	0.42	0.28	5	0.70	0.46	4	0.95	0.66

n = number of determinations

S.D. = standard deviation

Table B-11. Mean characteristics of the influent to each RBC unit when operating at 15°C.

Unit	A			B			C			D		
Parameter	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
Flow rate ℓ/d	8	276.3	7.6	8	285.3	5.4	11	282.0	6.8	12	288.6	7.2
COD Total mg/ℓ	8	79.3	25.0	8	144.5	49.7	11	192.6	72.8	12	265.2	96.8
COD Filtered mg/ℓ	5	25.5	11.7	6	48.0	18.7	9	64.0	26.9	9	87.1	37.2
SS - mg/ℓ	5	28.9	7.7	6	50.4	15.6	9	67.6	24.9	9	91.9	33.7
VSS - mg/ℓ	5	26.3	4.8	6	45.5	11.5	9	62.7	19.4	9	85.2	26.1
TKN - mg/ℓ	2	10.4	1.3	2	20.0	2.1	2	30.4	2.7	2	40.5	3.5
Ammonia-N, mg/ℓ	3	7.70	0.11	3	14.76	0.27	3	22.3	0.7	3	29.79	0.91
Nitrite-N, mg/ℓ	5	0.003	0.001	4	0.005	0.002	5	0.007	0.002	5	0.010	0.003
Nitrate-N, mg/ℓ	5	<0.04	<0.04	4	<0.04	<0.04	5	<0.04	<0.04	5	<0.04	<0.04

n = number of determinations

S.D. = standard deviation

Table B-12. Mean characteristics of the influent to each RBC unit when operating at 20°C.

Unit	A			B			C			D		
Parameter	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.	n	Mean	S.D.
Flow rate ℓ/d	12	280.2	3.7	10	283.7	3.7	12	287.4	6.5	12	292.2	4.8
COD Total mg/ℓ	12	145.5	45.7	10	202.3	66.0	12	256.7	85.2	12	281.9	95.2
COD Filtered mg/ℓ	3	44.9	13.8	3	62.6	18.5	3	79.1	25.6	4	83.2	29.3
SS - mg/ℓ	4	45.8	17.7	3	66.6	28.9	4	80.4	32.8	5	84.0	38.1
VSS - mg/ℓ	4	43.4	12.8	3	62.1	21.1	4	76.0	24.1	5	78.4	32.2
TKN - mg/ℓ	8	21.0	4.1	7	29.6	6.5	8	36.8	7.5	9	48.5	8.2
Ammonia-N, mg/ℓ	8	10.0	4.4	7	13.0	5.7	8	17.5	7.4	9	22.3	9.9
Nitrite-N, mg/ℓ	8	0.122	0.103	7	0.159	0.152	8	0.212	0.176	9	0.238	0.201
Nitrate-N, mg/ℓ	8	0.20	0.15	7	0.28	0.22	8	0.34	0.25	9	0.40	0.28

n = number of determinations

S.D. = standard deviation

Table B-13. Dissolved oxygen concentrations and pH values measured in various stages of RBC units when operating at 5°C.

Unit	Date	First Stage 6/18	7/5	7/9	Second Stage 6/18	7/5	7/9	Third Stage 6/18	7/5	7/9	Fourth Stage 6/18	7/5	7/9
B	DO - mg/ℓ	8.2	9.6	8.6	8.8	8.6	9.1	9.0	9.9	9.6	9.4	9.9	9.6
	pH	8.1	8.15	7.85	8.15	8.05	8.00	8.2	8.25	8.05	8.25	8.35	8.10
A	DO - mg/ℓ	6.8	8.3	7.8	8.3	9.1	8.6	8.7	9.5	9.0	9.3	9.9	9.1
	pH	8.00	8.05	7.85	8.00	8.25	8.00	8.00	8.30	8.10	8.15	8.35	8.15
C	DO - mg/ℓ	7.0	8.4	8.3	8.0	9.2	9.1	8.8	9.9	9.4	9.1	9.9	9.1
	pH	8.05	8.05	7.90	8.10	8.25	8.10	8.10	8.35	8.15	8.20	8.35	8.20
D	DO - mg/ℓ	6.7	7.6	7.8	8.1	8.3	8.6	8.3	8.6	8.7	8.7	9.2	8.8
	pH	8.10	8.00	8.00	8.20	8.15	8.15	8.20	8.25	8.20	8.25	8.35	8.20

Table B-14. Dissolved oxygen concentrations and pH values measured in various stages of RBC units when operating at 15°C.

Unit	Date	First Stage 12/1	12/24	Second Stage 12/1	12/24	Third Stage 12/1	12/24	Fourth Stage 12/1	12/24
A	DO - mg/ℓ	5.0		6.7		7.4		7.8	
	pH	7.80		7.85		7.90		8.00	
B	DO - mg/ℓ	3.9		5.3		6.8		7.5	
	pH	7.70		7.80		7.85		7.90	
C	DO - mg/ℓ	3.1	4.0	4.6	5.1	5.8	4.3	7.6	5.2
	pH	7.70	7.70	7.80	7.65	7.85	7.60	7.85	7.60
D	DO - mg/ℓ	2.5	2.8	3.6	3.2	4.5	3.4	10.2	3.9
	pH	7.65	7.80	7.80	7.75	7.85	7.65	7.85	7.50

Table B-15. Dissolved oxygen concentrations and pH values measured in various stages of RBC units when operating at 20°C.

Unit	Date	First Stage 3/9	4/8	Second Stage 3/9	4/8	Third Stage 3/9	4/8	Fourth Stage 3/9	4/8
A	DO - mg/ℓ	4.1	3.1	4.2	5.0	6.5	6.0	7.3	6.5
	pH	8.05	7.85	7.90	7.90	8.05	8.05	8.15	8.10
B	DO - mg/ℓ	3.5	2.2	3.7	4.3	4.0	5.7	6.4	6.5
	pH	8.10	7.85	7.90	7.90	7.85	8.05	8.00	8.15
C	DO - mg/ℓ	2.5	2.2	3.2	2.8	4.1	5.5	6.5	6.5
	pH	8.05	7.85	7.95	7.75	7.90	7.85	8.05	8.05
D	DO - mg/ℓ	2.1	1.7	2.8	2.3	3.4	5.4	6.1	5.5
	pH	7.75	7.85	7.80	7.80	7.75	7.90	7.95	8.00

Table B-16. Characteristics of the effluent from the four stages of RBC Unit A when operating at 5°C.

RBC Data T = 5°C Unit - A								
Parameter	5/29	6/6	6/20	6/26	7/1	7/3	7/5	7/7
Total COD - mg/ℓ								
First Stage	59.4	67.8	60.8		88.6	64.9	64.9	74.1
Second Stage	51.1	85.9	137.3		77.6	73.3	60.6	92.1
Third Stage	54.7	71.9	62.3		84.2	66.3	68.0	69.6
Fourth Stage	53.9	67.2	62.1		80.0	69.6	78.5	64.8
Filtered COD - mg/ℓ								
First Stage	25.3	--	27.5		28.6	26.1	21.4	25.0
Second Stage	--	44.1	39.6		32.3	21.9	20.8	23.6
Third Stage	24.0	24.7	25.3		24.4	21.7	20.1	24.1
Fourth Stage	25.5	24.0	24.7		30.0	21.0	18.5	23.4
SS - mg/ℓ								
First Stage	33	34	36		54	34	41	40
Second Stage	29	49	44		53	33	39	50
Third Stage	22	51	36		55	39	41	44
Fourth Stage	27	43	44		54	41	38	35
VSS - mg/ℓ								
First Stage	30	34	35		51	34	38	40
Second Stage	27	48	41		47	33	39	47
Third Stage	22	49	34		50	39	41	44
Fourth Stage	27	43	41		43	37	38	35
Filtered Kjeldahl-N mg/ℓ								
First Stage				28.6	21.1			25.3
Second Stage				33.6	20.9			25.2
Third Stage				27.6	29.6			25.8
Fourth Stage				50.0	20.9			24.1
Ammonia-N mg/ℓ								
First Stage				10.57	13.36			16.53
Second Stage				14.92	14.05			16.68
Third Stage				15.00	21.08			15.11
Fourth Stage				14.92	13.90			16.45
Nitrite - N mg/ℓ								
First Stage				0.142	0.295			0.395
Second Stage				0.168	0.420			0.500
Third Stage				0.192	0.420			0.520
Fourth Stage				0.230	0.475			0.490
Nitrate-N mg/ℓ								
First Stage				0.92	0.56			0.75
Second Stage				0.93	0.53			0.85
Third Stage				1.01	0.63			0.98
Fourth Stage				1.17	0.62			1.06

Table B-17. Characteristics of the effluent from the four stages of RBC Unit B when operating at 5°C.

Parameter	RBC Data T = 5°C Unit - B							
	5/28	6/4	6/18	6/25	6/30	7/4	7/6	7/9
Total COD - mg/ℓ								
First Stage	53.4	59.9	73.7		72.7	43.6	49.4	48.2
Second Stage	47.6	75.5	161.0		57.6	61.3	68.0	53.2
Third Stage	46.3	71.0	76.3		51.3	48.7	69.1	44.8
Fourth Stage	46.1	57.2	91.2		55.1	63.5	95.8	53.4
Filtered COD - mg/ℓ								
First Stage	18.5	24.5	27.3		24.5	25.0	14.8	18.7
Second Stage	18.5	21.6	39.1		18.2	22.8	12.1	16.4
Third Stage	18.9	20.9	21.3		16.4	15.4	12.7	13.7
Fourth Stage	18.7	19.5	32.3		24.9	14.1	13.0	16.0
SS - mg/ℓ								
First Stage	59.0	47.0	62.0		39.0	20.0	27.0	25.0
Second Stage	39.0	55.0	121.0		30.0	20.0	34.0	31.0
Third Stage	40.0	59.0	50.0		30.0	25.0	44.0	24.0
Fourth Stage	41.0	--	92.0		28.0	37.0	41.0	35.0
VSS - mg/ℓ								
First Stage	46.0	43.0	54.0		39.0	20.0	27.0	25.0
Second Stage	36.0	49.0	94.0		30.0	20.0	34.0	31.0
Third Stage	39.0	50.0	46.0		30.0	25.0	41.0	24.0
Fourth Stage	30.0	--	77.0		28.0	34.0	38.0	35.0
Filtered Kjeldahl-N mg/ℓ								
First Stage				40.6	16.3		19.6	
Second Stage				22.4	17.0		14.8	
Third Stage				33.6	16.6		19.6	
Fourth Stage				20.6	16.9		17.4	
Ammonia-N mg/ℓ								
First Stage				9.72	9.82		11.99	
Second Stage				9.72	9.57		9.53	
Third Stage				9.56	7.12		10.65	
Fourth Stage				9.64	6.99		11.02	
Nitrite-N mg/ℓ								
First Stage				0.134	0.270		0.188	
Second Stage				0.146	0.290		0.192	
Third Stage				0.158	0.295		0.168	
Fourth Stage				0.198	0.275		0.156	
Nitrate-N mg/ℓ								
First Stage				0.89	0.98		1.61	
Second Stage				0.95	1.06		1.83	
Third Stage				1.04	1.31		2.25	
Fourth Stage				1.22	1.38		2.34	

Table B-18. Characteristics of the effluent from the four stages of RBC Unit C when operating at 5°C.

RBC Data T = 5°C Unit - C								
Parameter	5/29	6/6	6/20	6/26	7/1	7/3	7/5	7/7
Total COD - mg/ℓ								
First Stage	149.1	186.6	120.5	104.0	152.9	155.9	132.6	101.2
Second Stage	141.7	--	102.3	99.9	130.3	123.2	130.2	81.1
Third Stage	143.6	151.1	111.4	116.4	112.3	109.2	141.9	93.1
Fourth Stage	104.9	--	120.5	124.7	137.1	127.9	139.6	82.0
Filtered COD - mg/ℓ								
First Stage	36.5	34.2	33.8	26.9	33.0	27.0	24.1	30.9
Second Stage	34.5	29.9	30.0	31.7	33.9	27.3	30.4	30.0
Third Stage	--	30.6	30.2	24.3	28.8	24.3	23.5	35.2
Fourth Stage	31.1	35.8	32.8	23.6	26.7	25.6	22.8	26.3
SS - mg/ℓ								
First Stage	78	150	60	44	98	41	38	48
Second Stage	89	79	74	61	69	31	38	45
Third Stage	99	92	60	95	75	28	46	55
Fourth Stage	55	87	77	89	84	36	56	44
VSS - mg/ℓ								
First Stage	69	118	55	44	88	41	38	48
Second Stage	80	67	62	52	57	31	38	42
Third Stage	76	82	55	75	65	28	46	53
Fourth Stage	55	73	62	73	72	36	52	44
Filtered Kjeldahl-N mg/ℓ								
First Stage				56.6	22.9			30.3
Second Stage				30.6	24.3			31.6
Third Stage				31.9	23.7			29.5
Fourth Stage				53.3	24.0			27.4
Ammonia-N mg/ℓ								
First Stage				17.66	14.98			16.45
Second Stage				18.15	15.52			15.70
Third Stage				16.77	14.83			17.20
Fourth Stage				21.05	15.52			16.15
Nitrite-N mg/ℓ								
First Stage				0.172	0.220			0.235
Second Stage				0.140	0.295			0.260
Third Stage				0.148	0.210			0.280
Fourth Stage				0.154	0.340			0.310
Nitrate-N mg/ℓ								
First Stage				0.91	0.53			0.46
Second Stage				1.00	0.41			0.69
Third Stage				1.01	--			0.72
Fourth Stage				0.99	0.56			0.94

Table B-19. Characteristics of the effluent from the four stages of RBC Unit D when operating at 5°C.

RBC Data T = 5°C Unit D

Parameter	5/28	6/4	6/18	6/23	6/25	6/30	7/2	7/4	7/6
Total COD - mg/ℓ									
First Stage	143.6	182.4	152.3	153.6		216.3	181.6	127.9	140.4
Second Stage	189.6	149.0	159.1	139.1		186.1	141.9	130.2	130.8
Third Stage	92.0	326.8	152.3	112.3		201.5	174.6	137.2	119.6
Fourth Stage	139.9	--	--	149.5		194.9	195.7	158.0	101.8
Filtered COD - mg/ℓ									
First Stage	45.3	42.8	43.6	40.5		44.5	34.7	36.0	28.7
Second Stage	38.6	41.2	40.2	60.4		40.0	36.7	33.7	28.8
Third Stage	41.0	40.5	36.4	33.9		37.6	32.1	31.3	31.3
Fourth Stage	39.3	37.6	51.2	--		32.4	29.3	30.4	32.7
SS - mg/ℓ									
First Stage	104	107	90	100		93	75	48	104
Second Stage	80	122	67	77		79	63	44	80
Third Stage	93	208	71	73		76	62	54	73
Fourth Stage	89	--	69	68		68	72	64	75
VSS - mg/ℓ									
First Stage	87	92	78	87		89	65	48	92
Second Stage	71	108	65	62		75	54	41	75
Third Stage	74	160	68	64		74	57	48	65
Fourth Stage	79	--	58	61		68	58	53	63
Filtered Kjeldahl-N mg/ℓ									
First Stage					32.0	29.7			47.6
Second Stage					39.9	29.3			38.1
Third Stage					25.0	29.6			37.9
Fourth Stage					40.9	28.3			31.8
Ammonia-N mg/ℓ									
First Stage					20.00	14.97			24.53
Second Stage					20.72	14.36			25.02
Third Stage					20.00	14.48			25.32
Fourth Stage					19.36	14.36			25.02
Nitrite-N mg/ℓ									
First Stage					0.156	0.240			0.194
Second Stage					0.160	0.265			0.240
Third Stage					0.166	0.315			0.275
Fourth Stage					0.178	0.350			0.305
Nitrate-N mg/ℓ									
First Stage					1.22	0.76			0.91
Second Stage					1.18	0.68			0.86
Third Stage					1.15	0.68			0.88
Fourth Stage					1.12	0.74			0.84

Table B-20. Characteristics of the effluent from the four stages of RBC Unit A when operating at 15°C.

RBC Data T = 15°C Unit - A

Parameter	12/10	12/13	12/14	12/19
Total COD - mg/ℓ				
First Stage	--	47.4	--	36.3
Second Stage	69.4	72.1	58.9	--
Third Stage	--	31.4	--	38.5
Fourth Stage	--	--	--	27.0
Filtered COD - mg/ℓ				
First Stage	23.2	24.4	19.2	17.2
Second Stage	--	18.6	25.1	--
Third Stage	--	20.5	--	20.6
Fourth Stage	--	21.5	--	21.1
SS - mg/ℓ				
First Stage	22	48	36	20
Second Stage	32	42	78	12
Third Stage	20	34	40	20
Fourth Stage	20	36	54	8
VSS - mg/ℓ				
First Stage	22	48	30	20
Second Stage	32	32	58	12
Third Stage	20	34	40	20
Fourth Stage	20	36	40	8
TKN - mg/ℓ				
First Stage		6		3
Second Stage		7		2
Third Stage		6		1
Fourth Stage		8		1
Ammonia-N mg/ℓ				
First Stage		1.79		0.31
Second Stage		0.87		0.30
Third Stage		0.75		0.12
Fourth Stage		0.67		0.12
Nitrite-N mg/ℓ				
First Stage		0.725	0.500	0.450
Second Stage		0.150	0.027	0.145
Third Stage		0.056	0.018	0.070
Fourth Stage		0.029	0.021	0.040
Nitrate-N mg/ℓ				
First Stage		9.02	7.75	7.05
Second Stage		10.85	7.72	7.30
Third Stage		10.69	9.23	7.58
Fourth Stage		10.97	9.98	7.76

Table B-21. Characteristics of the effluent from the four stages of RBC Unit B when operating at 15°C.

RBC Data T = 15°C Unit - B				
Parameter	12/12	12/18	12/20	12/21
Total COD - mg/ℓ				
First Stage	--	--	80.0	105.0
Second Stage	49.0	--	41.5	50.0
Third Stage	--	--	49.8	61.8
Fourth Stage	--	--	41.9	41.0
Filtered COD - mg/ℓ				
First Stage	--	23.0	--	33.6
Second Stage	22.9	--	20.9	25.7
Third Stage	24.2	--	24.1	--
Fourth Stage	25.5	--	19.2	19.7
SS - mg/ℓ				
First Stage	88	46	33	31
Second Stage	66	32	24	26
Third Stage	44	22	26	15
Fourth Stage	38	38	37	10
VSS - mg/ℓ				
First Stage	74	46	33	31
Second Stage	58	32	23	26
Third Stage	44	22	24	15
Fourth Stage	38	38	32	10
TKN - mg/ℓ				
First Stage	--	16.0		
Second Stage	--	9.0		
Third Stage	--	5.0		
Fourth Stage	--	4.0		
Ammonia-N mg/ℓ				
First Stage	19.15	10.00		
Second Stage	7.43	0.57		
Third Stage	3.93	0.12		
Fourth Stage	--	0.36		
Nitrite-N mg/ℓ				
First Stage	1.50	0.925		
Second Stage	0.925	0.475		
Third Stage	0.250	0.115		
Fourth Stage	0.125	0.043		
Nitrate-N mg/ℓ				
First Stage	0.75	3.32		
Second Stage	10.57	12.02		
Third Stage	11.50	13.13		
Fourth Stage	11.87	13.96		

Table B-22. Characteristics of the effluent from the four stages of RBC Unit C when operating at 15°C.

RBC Data T = 15°C Unit - C

Parameter	12/11	12/12	12/18	12/20	12/21	12/24	12/25
Total COD - mg/ℓ							
First Stage	--	159.0		96.0	161.0	91.3	75.5
Second Stage	--	69.0		131.0	81.0	104.0	91.4
Third Stage	--	130.0		61.0	55.3	70.8	133.0
Fourth Stage	--	--		47.5	42.9	133.0	84.8
Filtered COD - mg/ℓ							
First Stage	34.0	41.9		38.7	34.8	30.5	49.8
Second Stage	30.8	--		50.4	62.2	24.4	30.8
Third Stage	--	34.1		38.3	36.1	24.6	33.5
Fourth Stage	--	22.0		23.9	--	24.5	31.2
SS - mg/ℓ							
First Stage	208	96		44	68	115	88
Second Stage	122	94		43	40	60	59
Third Stage	182	114		36	45	85	76
Fourth Stage	114	72		23	27	93	90
VSS - mg/ℓ							
First Stage	174	86		40	63	103	73
Second Stage	108	84		37	40	60	55
Third Stage	156	92		36	45	59	68
Fourth Stage	100	72		23	27	77	75
TKN - mg/ℓ							
First Stage			26.0				
Second Stage			17.0				
Third Stage			8.0				
Fourth Stage			4.0				
Ammonia-N mg/ℓ							
First Stage		17.82	18.47				
Second Stage		--	9.64				
Third Stage		1.12	1.65				
Fourth Stage		0.66	0.20				
Nitrite-N mg/ℓ							
First Stage		0.80	1.38				
Second Stage		5.80	1.05				
Third Stage		1.75	0.68				
Fourth Stage		0.42	0.25				
Nitrate-N mg/ℓ							
First Stage		0.05	0.07				
Second Stage		10.20	9.95				
Third Stage		13.75	17.82				
Fourth Stage		16.57	20.50				

Table B-23. Characteristics of the effluent from the four stages of RBC Unit D when operating at 15°C.

RBC Data T = 15°C Unit - D

Parameter	12/10	12/13	12/14	12/18	12/19	12/24	12/25
Total COD - mg/ℓ							
First Stage	200.0	112.0	122.0		221.0	181.0	154.0
Second Stage	161.0	106.0	--		158.0	107.0	96.3
Third Stage	160.0	80.0	--		173.0	173.0	90.2
Fourth Stage	99.7	--	--		154.0	88.4	50.4
Filtered COD - mg/ℓ							
First Stage	61.0	50.3	46.7	--	57.6	41.4	41.4
Second Stage	40.0	42.7	40.9	--	35.9	27.1	32.5
Third Stage	74.3	58.5	--	39.0	68.1	24.3	31.7
Fourth Stage	49.9	39.9	33.2	33.7	50.2	--	32.4
SS - mg/ℓ							
First Stage	134	174	164	92	138	277	222
Second Stage	150	176	104	120	116	169	123
Third Stage	126	118	110	90	80	148	123
Fourth Stage	66	66	56	98	80	122	84
VSS - mg/ℓ							
First Stage	126	164	132	92	106	199	185
Second Stage	132	152	90	116	106	138	114
Third Stage	100	100	98	90	68	128	109
Fourth Stage	66	62	56	92	64	106	76
TKN - mg/ℓ							
First Stage		53			37		
Second Stage		39			30		
Third Stage		34			12		
Fourth Stage		7			11		
Ammonia-N mg/ℓ							
First Stage		27.50			31.16		
Second Stage		24.17			23.13		
Third Stage		16.83			9.84		
Fourth Stage		5.96			1.81		
Nitrite-N mg/ℓ							
First Stage		0.175	0.043		0.200		
Second Stage		3.800	2.425		2.250		
Third Stage		2.175	1.775		1.175		
Fourth Stage		1.325	1.275		0.850		
Nitrate-N mg/ℓ							
First Stage		0.26	0.32		0.23		
Second Stage		--	1.32		4.25		
Third Stage		7.82	9.22		13.82		
Fourth Stage		18.67	19.72		24.40		

Table B-24. Characteristics of the effluent from the four stages of RBC Unit A when operating at 20°C.

RBC Data T = 20°C Unit - A

Parameter	3/26	4/3	4/8	4/10	4/12
Total COD - mg/ℓ					
First Stage	--	88.7	198.4	116.1	82.1
Second Stage	52.7	56.7	116.1	105.8	101.0
Third Stage	74.9	121.1	80.0	110.8	74.2
Fourth Stage	78.9	105.1	81.3	97.4	80.5
Filtered COD - mg/ℓ					
First Stage	25.5	34.9	24.0	21.3	20.4
Second Stage	19.8	24.2	19.9	27.6	16.6
Third Stage	47.6	48.8	19.2	22.0	17.7
Fourth Stage	--	42.5	18.0	19.9	21.1
SS - mg/ℓ					
First Stage	38	62	87	82	53
Second Stage	41	44	98	87	60
Third Stage	37	65	61	93	61
Fourth Stage	36	54	58	78	60
VSS - mg/ℓ					
First Stage	34	53	70	65	53
Second Stage	41	40	77	69	53
Third Stage	37	62	51	77	54
Fourth Stage	36	52	52	63	52
Filtered Kjeldahl - mg/ℓ					
First Stage	10.87	--	13.15	11.15	
Second Stage	2.21	1.05	1.85	0.57	
Third Stage	1.70	0.25	1.43	0.42	
Fourth Stage	1.53	0.20	0.72	0.58	
Ammonia-N mg/ℓ					
First Stage	10.87	2.85	3.13	2.48	
Second Stage	0.28	0.17	0.35	0.37	
Third Stage	0.35	0.07	0.11	0.11	
Fourth Stage	0.13	0.04	0.11	0.09	
Nitrite-N mg/ℓ					
First Stage	0.525	0.825	1.080	1.740	
Second Stage	0.400	0.420	0.820	1.070	
Third Stage	0.275	0.230	0.375	0.395	
Fourth Stage	0.125	0.105	0.215	0.265	
Nitrate-N mg/ℓ					
First Stage	3.98	6.68	4.32	4.66	
Second Stage	12.60	7.88	6.18	8.63	
Third Stage	13.98	8.62	6.88	8.56	
Fourth Stage	13.88	8.14	8.14	9.38	

Table B-25. Characteristics of the effluent from the four stages of RBC Unit B when operating at 20°C.

RBC Data T = 20°C Unit - B					
Parameter	3/27	4/2	4/7	4/9	4/11
Total COD - mg/ℓ					
First Stage	77.4	130.6	91.1	86.5	104.5
Second Stage	--	43.1	127.0	88.7	93.0
Third Stage	--	52.6	90.5	75.7	75.3
Fourth Stage	143.0	157.6	--	120.2	72.0
Filtered COD - mg/ℓ					
First Stage	28.0	23.5	30.1	25.7	--
Second Stage	21.4	34.9	42.2	46.8	--
Third Stage	28.2	21.0	42.6	37.2	--
Fourth Stage	47.8	21.0	26.0	25.3	18.8
SS - mg/ℓ					
First Stage	66	36	64	75	68
Second Stage	45	54	81	60	47
Third Stage	82	56	66	45	42
Fourth Stage	90	55	55	84	53
VSS - mg/ℓ					
First Stage	59	36	60	73	65
Second Stage	40	52	57	57	47
Third Stage	72	46	58	44	42
Fourth Stage	71	54	54	77	51
Filtered Kjeldahl - mg/ℓ					
First Stage	28.90	5.61	6.74	17.00	
Second Stage	7.44	0.61	2.03	3.77	
Third Stage	1.62	0.72	0.75	1.89	
Fourth Stage	0.87	0.20	0.69	1.25	
Ammonia-N mg/ℓ					
First Stage	21.00	4.82	6.22	6.72	
Second Stage	7.43	0.41	0.84	2.27	
Third Stage	0.41	0.25	0.22	0.85	
Fourth Stage	0.09	0.16	0.09	0.39	
Nitrite-N mg/ℓ					
First Stage	2.200	1.875	1.025	0.800	
Second Stage	1.925	0.375	0.580	1.350	
Third Stage	0.525	0.185	0.460	1.050	
Fourth Stage	0.250	0.175	0.300	0.750	
Nitrate-N mg/ℓ					
First Stage	0.80	4.12	3.48	2.45	
Second Stage	10.82	9.88	7.72	6.65	
Third Stage	19.72	9.32	8.69	8.70	
Fourth Stage	20.25	9.38	8.85	9.75	

Table B-26. Characteristics of the effluent from the four stages of RBC Unit C when operating at 20°C.

RBC Data T = 20°C Unit - C					
Parameter	3/26	4/3	4/8	4/10	4/12
Total COD - mg/ℓ					
First Stage	94.5	96.4	98.8	98.9	143.0
Second Stage	154.4	85.6	114.6	187	174
Third Stage	171.5	45.6	97.5	129.5	97
Fourth Stage	--	84.9	96.0	111.8	196.0
Filtered COD - mg/ℓ					
First Stage	38.9	--	37.3	38.2	--
Second Stage	30.3	--	30.7	--	27.9
Third Stage	--	25.2	29.5	28.3	22.2
Fourth Stage	29.3	--	28.1	30.9	23.1
SS - mg/ℓ					
First Stage	89	62	77	87	123
Second Stage	74	53	77	137	112
Third Stage	110	58	76	78	90
Fourth Stage	124	66	80	103	83
VSS - mg/ℓ					
First Stage	78	62	66	78	107
Second Stage	67	52	70	107	99
Third Stage	91	54	67	63	74
Fourth Stage	104	64	67	86	70
Filtered Kjeldahl-N mg/ℓ					
First Stage	35.00	14.55	25.20	20.40	
Second Stage	17.70	2.32	10.65	9.73	
Third Stage	4.49	2.60	3.38	1.60	
Fourth Stage	2.27	1.65	1.62	1.01	
Ammonia-N mg/ℓ					
First Stage	25.73	14.55	10.88	16.88	
Second Stage	11.35	0.91	1.28	2.79	
Third Stage	2.96	0.27	0.20	0.47	
Fourth Stage	0.26	0.11	0.06	0.08	
Nitrite-N mg/ℓ					
First Stage	0.345	0.850	0.275	0.725	
Second Stage	0.950	1.200	1.020	1.700	
Third Stage	0.900	0.350	0.480	0.930	
Fourth Stage	0.375	0.150	0.250	0.560	
Nitrate-N mg/ℓ					
First Stage	0.71	1.40	2.22	1.02	
Second Stage	10.80	12.55	9.38	10.30	
Third Stage	--	14.90	11.62	13.97	
Fourth Stage	--	14.60	11.95	11.94	

Table B-27. Characteristics of the effluent from the four stages of RBC Unit D when operating at 20°C.

RBC Data T = 20°C Unit - D						
Parameter	3/20	3/27	4/2	4/7	4/9	4/11
Total COD - mg/ℓ						
First Stage	286.7	145.9	130.5	229.4	147.7	216.0
Second Stage	--	228.5	123.1	131.8	172.6	192.0
Third Stage	--	103.1	115.8	97.7	111.4	174.0
Fourth Stage	--	257.0	140.2	152.2	140.9	240.0
Filtered COD - mg/ℓ						
First Stage	35.2	40.7	42.4	46.5	40.9	36.9
Second Stage	25.6	--	30.9	26.2	31.9	26.1
Third Stage	39.4	--	25.0	29.3	32.8	--
Fourth Stage	36.7	--	23.8	31.1	26.4	22.0
SS - mg/ℓ						
First Stage	234	73	68	83	56	183
Second Stage	226	190	98	98	56	89
Third Stage	194	95	109	66	57	89
Fourth Stage	184	116	102	88	103	101
VSS - mg/ℓ						
First Stage	174	70	62	71	52	158
Second Stage	184	149	88	83	52	80
Third Stage	162	84	92	61	53	82
Fourth Stage	148	97	91	77	88	89
Filtered Kjeldahl-N mg/ℓ						
First Stage	--	45.10	15.45	18.60	32.80	
Second Stage	--	20.34	2.76	1.75	17.00	
Third Stage	--	5.59	1.40	2.64	6.12	
Fourth Stage	--	2.73	1.34	4.82	2.72	
Ammonia-N mg/ℓ						
First Stage	32.38	37.0	13.31	15.38	15.86	
Second Stage	14.81	20.34	2.76	0.33	6.61	
Third Stage	2.32	4.71	0.19	0.69	3.15	
Fourth Stage	0.44	0.45	0.25	3.18	1.06	
Nitrite-N mg/ℓ						
First Stage	0.770	0.290	1.125	0.350	0.225	
Second Stage	1.350	0.950	1.175	0.325	1.275	
Third Stage	0.800	1.000	0.225	0.800	1.300	
Fourth Stage	0.150	0.375	0.100	1.175	0.670	
Nitrate-N mg/ℓ						
First Stage	0.33	0.51	0.88	2.40	2.78	
Second Stage	16.15	11.55	12.32	13.18	7.48	
Third Stage	31.20	25.00	16.78	12.95	11.45	
Fourth Stage	33.80	29.12	16.65	9.82	14.43	

Table B-28. Total organic carbon concentration (mg/ℓ) in RBC Units A and C when operating at 20°C.

Date	3/13		4/3		4/8	
Unit	Unfiltered	Filtered	Unfiltered	Filtered	Unfiltered	Filtered
A						
First Stage	15.7	13.8	25.6	16.8	46.2	14.2
Second Stage	9.8	8.4	11.5	11.0		8.7
Third Stage	10.4	7.9	13.1	8.6		
Fourth Stage	16.9	8.5	16.8	8.2		
C						
First Stage	27.0	17.0	33.9	17.9	32.4	17.7
Second Stage	23.2	20.7	24.9	13.8		16.3
Third Stage	42.7	21.5	34.0	13.9		
Fourth Stage	22.3	19.6	20.8	12.8		

Table B-29. Total organic carbon concentration (mg/ℓ) in RBC Units B and D when operating at 20°C.

Date	3/20		4/2		4/7	
Unit	Total	Filtered	Total	Filtered	Total	Filtered
B						
First Stage	20.6	12.5	24.2	15.8		14.1
Second Stage	14.9	13.6	24.2	14.5		9.4
Third Stage	14.5	8.6	22.6	15.2		
Fourth Stage	16.8	8.0	31.9	14.8		
D						
First Stage	23.8	17.8	30.6	23.1	35.1	21.2
Second Stage	41.2	14.0	38.6	17.6		13.2
Third Stage	26.5	13.4	33.9	14.6		
Fourth Stage	39.7	12.4	36.1	12.1		

Table B-30. Mean influent concentrations of nitrogen forms when operating at 20°C for the period 4/2 to 4/12.

Parameter	A	B	C	D
TKN-mg/ℓ				
n	5	5	5	5
Average	20.94	26.69	37.09	39.00
S.D.	2.75	4.59	5.34	7.05
Ammonia-N mg/ℓ				
n	5	5	5	5
Average	7.64	10.82	13.51	15.76
S.D.	1.24	1.80	2.11	2.48
Nitrite-N mg/ℓ				
n	5	5	5	5
Average	0.117	0.188	0.205	0.273
S.D.	0.115	0.170	0.198	0.246
Nitrate-N mg/ℓ				
n	5	5	5	5
Average	0.20	0.36	0.35	0.52
S.D.	0.16	0.21	0.29	0.31
Flow rate-ℓ/d				
n	6	6	6	6
Average	279.6	283.2	289.2	294.0
S.D.	2.9	3.7	7.1	2.9

n = number of determinations
S.D. = standard deviation

Table B-31. Mass of attached biomass on the experimental RBC units, grams.

Temperature °C		5		15		20	
Unit	Stage	Total Solids	Volatile Solids	Total Solids	Volatile Solids	Total Solids	Volatile Solids
A	1	37.7	26.8	41.3	29.1	39.4	26.8
	2	22.9	14.9	11.0	6.9	7.8	5.0
	3	25.0	16.3	5.4	4.3	4.4	2.2
	4	17.8	11.1	3.4	2.2	2.5	1.6
B	1	35.4	23.8	63.1	38.0	89.5	59.7
	2	17.3	11.3	56.5	30.9	12.3	8.3
	3	16.3	10.4	19.5	11.4	3.4	2.2
	4	13.6	8.7	12.0	6.5	2.0	1.2
C	1	55.9	36.7	76.1	49.7	81.9	54.9
	2	32.1	20.7	56.9	37.7	37.5	24.2
	3	17.0	10.7	37.8	22.5	13.2	8.6
	4	25.7	15.1	24.7	14.4	7.8	4.6
D	1	53.8	34.6	59.5	48.8	88.9	58.7
	2	56.4	35.8	61.0	40.6	47.6	35.1
	3	42.3	27.9	33.0	24.2	7.3	4.6
	4	33.2	21.4	31.0	17.6	8.9	6.6

Table B-32. Percentage of volatile solids in attached biomass.

Temperature °C		5	15	20
Unit	Stage			
A	1	71.11	70.42	68.10
	2	65.19	62.60	64.00
	3	65.00	79.28	48.90
	4	62.47	64.43	63.20
B	1	67.34	60.25	66.70
	2	65.41	54.71	67.50
	3	63.88	58.31	64.70
	4	63.63	53.83	61.80
C	1	65.70	65.35	67.00
	2	64.40	66.17	64.40
	3	62.81	59.49	64.80
	4	58.79	58.42	58.80
D	1	64.29	82.05	66.00
	2	63.48	66.53	73.70
	3	65.98	73.46	63.10
	4	64.33	55.96	67.70

APPENDIX C

KINETIC MODELS--DATA

Table C-1. Summary of the values used in equations to calculate the yields and decay rates.

Temperature °C	Unit	$\frac{Q(S_0-S_1)}{A\bar{X}}$	$\frac{QX_1}{A\bar{X}}$
5	B	0.76065	0.43347
	A	0.99168	0.40010
	C	0.89081	0.50148
	D	1.17094	0.69837
15	A	0.55355	0.28485
	B	0.87242	0.34536
	C	0.87551	0.50953
	D	1.27445	0.84806
20	A	1.25776	0.57504
	B	0.83399	0.27847
	C	1.14437	0.40938
	D	1.20215	0.48683

Q = Influent flow rate, m^3/d

S_0 = Influent total COD, mg/ℓ

S_1 = First stage effluent filtered COD, mg/ℓ

X_1 = First stage effluent VSS, mg/ℓ

$\bar{X}$ = Attached biomass in the first stages, g/m^2

A = First stages surface area, m^2

Table C-2. Summary of mean steady state data used to determine the kinetic constants for substrate removal in the first stages of the RBC process.

Temperature °C	Unit	Q m^3/d	S_0 mg/ℓ	S_1 mg/ℓ	$A\bar{X}$ grams	$Q(S_0-S_1)/A$ $g/m^2/d$	$Q(S_0-S_1)/A\bar{X}$ g/g/d	$Q(S_0/S_1-1)/A$ $m^3/m^2/d$
5	A	0.2867	118.4	25.7	26.8	18.03059	0.99168	0.70158
5	B	0.2842	85.6	21.9	23.8	12.28191	0.76065	0.56082
5	C	0.2940	142.0	30.8	36.7	22.17965	0.89081	0.72012
5	D	0.3028	173.3	39.5	34.6	27.48619	1.17094	0.69585
15	A	0.2763	79.3	21.0	29.1	11.71512	0.55355	0.55786
15	B	0.2853	144.5	28.3	38.0	24.11044	0.87242	0.85196
15	C	0.2820	192.6	38.3	49.7	31.64553	0.87551	0.82625
15	D	0.2886	265.2	49.7	48.8	45.23149	1.27445	0.91009
20	A	0.2802	145.5	25.2	26.8	22.86843	1.25776	0.90748
20	B	0.2837	202.3	26.8	59.7	33.77839	0.83399	1.26039
20	C	0.2874	256.7	38.1	54.9	42.62255	1.14437	1.11870
20	D	0.2922	281.9	40.4	58.7	47.87402	1.20215	1.18500

Q = Influent flow, m^3/d

S_0, S_1 = Influent and effluent COD for the first stages, mg/ℓ

A = Surface area/stage, m^2

$\bar{X}$ = Attached biomass in first stage, g/m^2

Table C-3. Summary of the results obtained with the models for carbonaceous substrate removal in the first stages of the RBC units.

	5°C			15°C			20°C		
Model	Correlation Coefficient	k, k_m, k_a	K_s, K_m	Correlation Coefficient	k, k_m, k_a	K_s, K_m	Correlation Coefficient	k, k_m, k_a	K_s, K_m
$Q(S_0-S_1) = \frac{kA\bar{X}S_1}{K_s+S_1}$	0.8505	2.495	47.3	0.9495	7.761	262.2	0.3378	1.513	12.5
$Q(S_0-S_1) = \frac{k_mAS_1^2}{K_m+S_1}$	0.7472	1.057	16.7	0.8864	1.798	42.5	0.4625	1.552	12.8
$Q(S_0-S_1) = \frac{k_aAS_1}{K_s+S_1}$	0.9603	-45.5	-97.3	0.9647	-36.3	-81.6	0.8899	-112	-134.6

Q = Flow rate, m^3/d

S_0, S_1 = Influent and effluent COD mg/ℓ

A = Total surface area/stage, m^2

$\bar{X}$ = Total attached biomass in first stage, gVS/m^2

k = Maximum reaction rate, 1/d

k_m = Maximum reaction rate, $g/m^2/mg/ℓ/d$

k_a = Maximum reaction rate, $g/m^2/d$

K_s = Half saturation constant, mg/ℓ COD

K_m = Constant, mg/ℓ COD

Table C-4. Summary of the experimental data used to develop the predictive model for the second through fourth stages of the RBC units.

Unit	Flow ℓ/d	Mean* Temperature °C	Filtered COD - mg/ℓ				
			First Stage	Second Stage	Third Stage	Fourth Stage	Mean*
C	0.2940	4.6	30.8	31.0	28.1	28.1	29.1
D	0.3028	4.2	39.5	40.0	35.5	36.1	37.2
B	0.2853	14.6	28.3	23.2	24.2	21.5	23.0
C	0.2820	14.5	38.3	39.7	33.3	25.4	32.8
D	0.2886	15.1	49.7	36.5	49.3	39.9	41.9
C	0.2874	19.9	38.1	29.6	26.3	27.9	27.9
D	0.2922	19.7	40.4	28.1	31.6	28.0	29.2

(*) calculated as mean values of stages 2, 3, and 4.

Table C-5. Summary of mean steady state ammonia-N data used in the linear regression analysis to determine the kinetic parameters.

Temperature °C	Unit - Stage	Influent Flow ℓ/d	Ammonia-N mg/ℓ		Ammonia-N Removal rate, $gN/m^2/d$
			Influent	Effluent	
15	A-1	276.3	7.70	1.05	1.336
15	B-2	285.3	14.58	4.00	2.195
15	C-3	282.0	9.64	1.39	1.712
15	D-3	288.6	23.65	13.34	2.164
15	D-4	288.6	13.34	3.89	1.984
20	B-2	283.2	5.92	1.17	0.913
20	D-2	294.0	14.85	3.23	2.318
20*	B-2	288.0	21.0	7.43	2.651
20*	B-3	288.0	7.43	1.62	1.135
20*	C-2	280.8	25.73	11.35	2.739
20*	C-3	280.8	11.35	2.96	1.598
20*	D-2	284.4	34.69	17.58	3.301
20*	D-3	284.4	17.58	3.52	2.713

(*) first period data.

Table C-6. Summary of mean steady state ammonia-N concentrations and removals.

Temperature					
15°C			20°C		
Unit-Stage	NH_4-N mg/ℓ	NH_4-N Removal rate-g/m^2/d	Unit-Stage	NH_4-N mg/ℓ	NH_4-N Removal rate-g/m^2/d
A-1	1.05	1.336	A-2	0.30	0.478
A-2	0.59	0.092	A-3	0.10	0.038
A-3	0.44	0.030	A-4	0.08	0.004
A-4	0.40	0.008	B-2	1.17	0.913
B-2	4.00	2.195	B-3	0.44	0.140
B-3*	2.03	0.409	B-4	0.21	0.044
B-4	0.36	0.347	C-2*	1.66	3.598
C-2*	9.64	1.745	C-3	0.61	0.390
C-3	1.39	1.712	C-4	0.05	0.075
C-4	0.43	0.199	D-2	3.23	2.318
D-3	13.34	2.164	B-2	7.43	2.651
D-4	3.89	1.984	B-3	1.62	1.135
			B-4	0.09	0.299
			C-2	11.35	2.739
			C-3	2.96	1.598
			C-4	0.26	0.514
			D-2	17.58	3.301
			D-3	3.52	2.713
			D-4	0.45	0.592

(*) were not considered in performing regression analysis.

Table C-7. Summary of calculated maximum ammonia-N removal rates (gN/m^2/d) in the first stages of the RBC units.

Temperature (°C)	15			20		
Unit	Observed	Simulated*	f_1	Observed	Simulated**	f_1
A	1.336	1.344	0.994	0.914	1.199	0.762
B	0.037	2.108	0.018	0.941	1.652	0.570
C	0.851	2.242	0.380	<0.00	2.011	<0.000
D	0.090	2.279	0.040	0.182	2.276	0.080

* The calculated values are based on Equation 82.

$$R = \frac{2.334\ (C-0.4)}{0.45+\ (C-0.4)} \qquad (82)$$

** The calculated values are based on Equation 83.

$$R = \frac{3.74\ C}{2.8\ +\ C} \qquad (83)$$

where

R = Ammonia-N removal rate, gN/m^2/d

C = Ammonia-N concentration, mg/ℓ

Table C-8. Summary of equations and order of application in estimation procedure to calculate ammonia-N concentrations.

Stage to be Estimated	Equations Applied
First Stage Ammonia-N	I. $\frac{Q(C_o-C_{1m})}{A_1} = \frac{k_N C_{1m}}{K_N+C_{1m}}$ II. $C_{1m} = \frac{-b+\sqrt{b^2-4ac}}{2a}$ $a = Q/A_1$ $b = k_N + \frac{Q}{A_1}K_N - \frac{Q}{A_1}C_o$ $c = \frac{-Q}{A_1}(C_o K_N)$ III. $\frac{Q}{A_1}(C_o-C_1) = f_1 \frac{Q}{A_1}(C_o-C_{1m})$ IV. $C_1 = f_1 C_{1m} + C_o(1-f_1)$
Second Stage Ammonia-N	V. $\frac{Q(C_1-C_2)}{A_2} = \frac{k_N C_2}{K_N+C_2}$ VI. $C_2 = \frac{-b+\sqrt{b^2-4ac}}{2a}$ $a = Q/A_2$ $b = k_N + \frac{Q}{A_2}K_N - \frac{Q}{A_2}C_1$ $c = \frac{-Q}{A_2}C_1 K_N$
Third Stage and Fourth Stage Ammonia-N $i = 3,4$	VII. $\frac{Q(C_{i-1}-C_i)}{A} = \frac{k_N C_i}{K_N+C_i}$ VIII. $C_i = \frac{-b+\sqrt{b^2-4ac}}{2}$ $a = Q/A_i$ $b = k_N + \frac{Q}{A_i}K_N - \frac{Q}{A_i}C_{i-1}$ $c = \frac{-Q}{A_i}K_N C_{i-1}$