**ENERGY CONSUMPTION AND PERFORMANCE OF
A ROTATING BIOLOGICAL CONTACTOR
TREATING HIGH-STRENGTH WASTEWATER**

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by

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A thesis
presented to Ryerson University
in partial fulfillment of the
requirements for the degree of
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ABSTRACT

Energy Consumption and Performance of a Rotating Biological Contactor Treating High-Strength Wastewater

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Clean water availability, energy costs and the environmental impact of energy usage are major concerns all over the world. At the same time, the Rotating Biological Contactor (RBC) has emerged as a low energy-consuming technology used in wastewater treatment which compares favorably with other treatment methods. RBC is a fixed-film bioreactor employing rotating discs to provide support medium for the microbial growth and to supply dissolved oxygen. RBCs, when applied in the treatment of high strength wastewater, demand some modifications such as the addition of aeration systems or change the flow configuration. Aeration systems certainly reduce the footprint but at the cost of energy consumption. Therefore, the optimization of energy consumption in a modified RBC is a very relevant research objective.

This thesis is an investigation on energy optimization in a commercial scale RBC modified with an aeration system and treating high strength synthetic wastewater. The coarse bubble diffuser was replaced by fine bubble air diffusers. To study energy consumption a mono-block main drive system and the central compressed air supply were replaced by a three phase motor with variable frequency drive and an aeration blower respectively. Removal performance and unit energy consumption were studied at various combinations of rotating speed (2.5-5 RPM) and rate of aeration (0-15 SCFM). Constant hydraulic ($0.017 \text{ m}^3/\text{m}^2\text{-day}$), organic ($86.1 \text{ gCOD}/\text{m}^2\text{-day}$) and ammonia ($3.444 \text{ gNH}_3\text{-N}/\text{m}^2\text{-day}$) loadings were maintained throughout the study. The modified RBC was able to remove 34 to 96% COD and 21 to 68% ammonia depending on the aeration rate and angular velocity. The suspended growth section of the modified RBC contributed 47 to 85% and 38 to 87% of the total removal of COD and ammonia respectively. Conversion of ammonia-nitrogen to nitrate-nitrogen was observed very negligible at 0.26 to 1.59%. The angular velocity, 3.66 RPM and the rate of aeration 8.13 SCFM, were found to be the optimum parameters to achieve minimum unit energy consumption of 1.31 KWH/kg CODr. A mathematical model correlating energy consumption per unit oxygen demand with the rate of aeration and the angular velocity was developed.

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ABBREVIATIONS AND ACRONYMS

ABC	Active Biological Contactors
AOTR	Actual Oxygen Transfer Rate
APHA.	American Public Health Association
ASCE	American Society of Civil Engineers
ASP	Activated Sludge Process
AWWA	American Water Works Association
BOD	Biological Oxygen Demand
BOD₅	5-Day Biological Oxygen Demand
BOD_r	Biological Oxygen Demand Removed
CF	Continuous Flow
COD	Chemical Oxygen Demand
COHNS	Symbolic Representation of Organic Matter in Wastewater
CSTR	Continuous Stirred Tank Reactor
DO	Dissolved Oxygen
DSM	Demand-Side Management
EA	Extended Aeration
EPRI	Electric Power Research Institute
ETFB	External Circulating Three-phase Fluidized bed Bioreactors
ETV	Environmental Technologies Verification

FBBR	Fluidized Bed Bioreactor
FRP	Fiber Reinforced Plastic
G7	Group of 7 leading industrial countries
GHG	Greenhouse Gas
HDPE	High-density polyethylene
HRT	Hydraulic Retention Time
IAMU	Iowa Association of Municipal Utilities
IF	Intermittent Flow
ISE	Ion Selective Electrode
LDO	Luminescent Dissolved Oxygen
LED	Light Emitting Diode
NEMA	National Electrical Manufacturers Association
NYSERDA	New York State Energy Research and Development Authority
ORP	Oxidation Reduction Potential
OUR	Oxygen Uptake Rate
POTW	Publicly Owned Treatment Works
PVC	Poly-Vinyl Chloride
RAS	Returned Activated Sludge
RBC	Rotating Biological Contactor
RPM	Revolutions per minute
SBOD	Soluble Biological Oxygen Demand

SBR	Sequencing Batch Reactor
SCFM	Standard Cubic Feet Per Minute
SOB	Sulphide Oxidizing Bacteria
SOTE	Standard oxygen-transfer Efficiency
SOTR	Standard Oxygen Transfer Rate
SRT	Solids Residence Time
TFB	Three-phase fluidized bed bioreactors
TOC	Total Organic Carbon
TOR	Theoretical Oxygen Requirement
UOD	Ultimate Oxygen Demand
USEPA	US Environmental Protection Agency
VFD	Variable Frequency Drive
WAS	Waste Activated Sludge
WEF	Water Environment Federation
WPCF	Water Pollution Control Facilities
WW	Wastewater
WWTP	Wastewater Treatment Plant

NOMENCLATURE

A	Total surface area of discs.....	m^2
A_e	Exposed area of the discs.....	m^2
A_i	Submerged area of the discs	m^2
$AOTR$	Actual oxygen transfer rate under field-operating conditions	$g\ O_2 / s$
C	Concentration of a reactant in the reactor at time t.....	g/m^3
C_0	Initial concentration of a reactant.....	g/m^3
C_s	Oxygen saturation concentration for clean water at field-operating conditions	mg/L
$C_{s(20)}$	Dissolved oxygen saturation concentration in clean water at 20 °C and 1 atm	mg/L
C_w	Operating oxygen concentration in wastewater.....	mg/L
d	Disc spacing	m
D	Diameter of the disc	m
D_m	Molecular diffusivity of oxygen	m^2/s
E	Enhancement of the oxygen transfer rate caused by the biomass	
F_{air}	Rate of aeration.....	$SCFM$
g	Acceleration due to gravity.....	m/s^2
g_c	Gravitational conversion factor	$kg\ m\ kg^{-1}\ min^{-2}$
H	Height of submergence of the disc from the bottom	m
K	First-order reaction rate constant.....	s^{-1}

K'_{La}	Overall biochemical oxygen mass-transfer coefficient in the reactor.....min ⁻¹
K_L	Gas-Liquid Mass-transfer coefficient for the liquid film exposed to air.....m/s
K_{La}	Gas-Liquid Volumetric Oxygen mass-transfer coefficient.....s ⁻¹
K_{La_w}	Overall physical oxygen mass-transfer coefficient in wastewater.....min ⁻¹
n	Number of stages in an RBC
N	Speed of rotations ⁻¹
N_p	Power number
p_1	Absolute inlet pressure..... atm
p_2	Absolute outlet pressure..... atm
P_{adb}	Power requirement for adiabatic compression..... kW
PW_{blower}	Actual wired power drawn by blower..... kW
$PW_{main\ drive}$	Actual wired power drawn by main drive..... kW
P_w	Requisite power..... kW
Q	Volumetric flow rate..... m ³ /s
r	Radius of the disc.....m
R	Universal Gas Constant, 8.314 kJ/kmol °K
r_c	Rate of reaction g/m ³ s
Re	Reynolds number
r_g	Rate of Biomass Growth mg VSS/L.day
r_o	Oxygen Uptake Ratemg O ₂ /L.day

r_{su}	Rate of Substrate Utilization mg bsCOD/L.day
$SOTR$	Standardized oxygen transfer rate under test conditions at 20°C and zero dissolved oxygen g O ₂ / s
t	Time s
T	Treatment temperature °C
T_1	Absolute inlet temperature °K
TOR	Theoretical oxygen requirement g O ₂ / s
V	Reactor volume m ³
V'_L	Effective volume of liquid in the reactor without biomass m ³
V_L	Effective volume of liquid in the reactor m ³
w	Mass flow rate of air kg/s

Greek Letters

α	Oxygen transfer correction factor, to estimate the K_La values for wastewater from values for clean water.
α'	Empirical constant, a product of the Reynolds number and power number
β	Salinity-surface tension correction factor, to correct the oxygen transfer rate for differences in oxygen solubility due to constituents in the wastewater
$\eta_{blower\ motor}$	Efficiency of Blower Motor
η_{blower}	Efficiency of Blower
$\eta_{main\ drive\ gear}$	Overall efficiency of main drive reducing gear system
$\eta_{main\ drive\ motor}$	Efficiency of main drive motor

λ_1	Frictional empirical constant..... kW min ² m ⁻⁴
λ_2	Frictional empirical constant..... kWmin ³ m ⁻⁵
μ	Viscosity of wastewater kg/m s
θ	Temperature Correction factor for the Volumetric Mass Transfer Coefficient $K_L a$
ρ	Density of wastewater kg/m ³
τ	Average Residence Time (V/Q) s
ω	Angular speed of rotation s ⁻¹

1. INTRODUCTION

1.1 Energy Use and Environmental Damage

Canada is the third highest energy user per capita (www.cleanairpartnership.org). Canada's high energy use is partly due to its cold winters and high standard of living. However, Denmark and Sweden experience similar weather conditions and use far less energy than we do. Germany, another country that has a high standard of living, also uses far less energy than we do. Indeed, these countries use 25-50% less energy per capita than Canada. (www.cleanairpartnership.org). Developed societies take energy for granted. Many of us may not even be aware of what our energy use is costing the natural environment, our health and the future of our children.

*"We do not inherit the Earth from our parents,
we borrow it from our children."*

- Saint Exupery

Energy generation brings with it emissions of harmful substances. The average Canadian produces over 5 tones of greenhouse gas (GHG) emissions per year. Canadians consume 72% more energy per person than the G7 average and our energy use is growing. Having only 0.5% of the world's population, we generate 3% of the world's greenhouse gases (GHGs).

1.2 Energy Consumption in Wastewater Treatment

Wastewater treatment systems reduce environmental pollution in receiving waters, but create other forms of pollution through energy consumption. This environmental impact is due to the Green House Gas (GHG) emissions associated with wastewater treatment including power generation, plus those generated from the degradation of pollutants. (Fact sheet, 2004)

Electricity is the largest expense in a wastewater treatment plant (WWTP) budget. Energy costs consume 15 to 30% of large WWTPs' and 30 to 40% of small WWTPs' operation and maintenance budgets (Kennedy, 1997). The California Energy Commission (1990) reported that WWTPs are often the single largest electricity users in a local government, accounting for 1 to 2% of a community's energy usage (USEPA, 1973) and 0.1 to 0.3% of the nation's total energy usage (Kennedy, 1997).

U.S. Publicly Owned Treatment Works (POTWs) consumed approximately 21 billion kWh of electricity in 2000. Privately operated wastewater treatment facilities are estimated to consume roughly twice that of POTWs. (EPRI, 2002) The New York State Energy Research and Development Authority (NYSERDA) estimated that municipal wastewater treatment plants (WWTPs) in New York State consume about 1.5 billion kWh of electricity each year (EPRI, 1993).

A survey carried out in Iowa (Sauer and Kimber, 2002) reported 236 million kWh energy was consumed to operate wastewater treatment systems in 1999-00, which cost 12 million dollars. These represented 0.3% of the total energy consumed in Iowa and 0.2% of the total electric costs for the year 2000 (IAMU, 2002). In terms of energy consumption per unit volume treated, wastewater treatment cost \$2.1/1,000 gal and consumed 1.57 kWh/1,000 gal for the whole WWTP which breaks down in 1.15 kWh/1,000 gal for treatment system (73% of total) and 0.42 kWh/1,000 gal for the collection system. (Sauer and Kimber, 2002)

Evans and Laughton (1994) estimated that Ontario consumes a total of 788 million kWh per year to treat wastewater at its 414 WWTPs. These plants ranged from plants with only primary treatment to plants with advanced tertiary treatment. Power consumption per unit volume treated was in the range 0.14 to 1.13 kWh/m³. Energy use may vary significantly for a given size wastewater treatment plant (WWTP), depending on its location, strength of wastewater, hydraulic conditions, level of treatment, in-plant recovery, type of treatment process selected, and mode of operation.

The secondary wastewater treatment process is usually the most energy intensive unit process at municipal wastewater treatment plants. Most often, aeration and pumping systems will be the largest energy users. A typical secondary treatment plant may use as much as 1500 to 1700 kWh of electricity to treat one million gallons of sewage and manage the resulting sludge and residuals (NYSERDA, 1993).

Energy costs are substantial at WWTPs and is an important part of wastewater treatment. Energy costs continue to steadily rise because of the cost of fuels, inflation, and increasingly stringent wastewater discharge requirements. It is generally believed that as long as there is a constant and seemingly unlimited supply of energy available, there is no need to think about saving energy, but that is not the case.

1.3 The New Energy Crisis and Demand Side Management

Many economists claim that despite the initial pain, the 1970s energy-efficiency measures made the United States economy more competitive. Manufacturing efficiency boosted then still shows up today. Even so, per capita energy use has been rising since the mid-1980s. "The country spent about 20 years

learning how to save energy and the next 12 years forgetting more than we ever learned" (Clayton, 2005)

The world is currently facing a new energy crisis. Since we are all part of the problem, we must be part of the solution too. We must act now to reduce our energy use and reduce the threat of global climate change and air pollution. This means that as Canadians, we have a responsibility to reduce our energy consumption and in doing so reduce our contribution to climate change.

The following scenario is all too familiar in the energy industry: The energy user demands more power, the utility company increases its supply to accommodate demand, and both parties end up spending more money and gobbling up more resources. A more cost-effective approach, however, may be to reduce the demand side of the equation, rather than automatically address the supply side. Managing the demand side of the energy is called 'demand-side management' (DSM). NYSERDA's Municipal Wastewater Treatment and Sludge Management Program define demand-side management as an opportunity a WWTP has to reduce total energy cost. The California legislature recognized, in 1988, that energy conservation opportunities exist at almost every WWTP. As such, the commission targeted WWTPs for energy assistance programs. Electric utilities in New York State have encouraged DSM to help control or lower energy costs and make energy available for new customers without constructing additional facilities. DSM includes energy conservation, energy efficiency, and control of power requirements.

1.4 Energy Management and Energy Efficiency

The world has become sensitive to "consumption" to the point where management of resources becomes not only economically viable but the right thing to do. The shortage of oil and the accompanying high cost of energy have led to a reevaluation of energy requirements in an attempt to locate potential areas for energy reduction.

There are two ways that we can reduce our demand for energy - through energy efficiency and energy management. Although both terms sometimes seem to mean the same thing, they actually have two separate definitions. **Energy efficiency** means using less energy to do the same job, which involves technologies and operational changes that reduce the amount of energy required for the job. On the other hand, **energy management** means avoiding the need to use energy, which involves behavioral and operational changes that reduce overall energy consumption. Both energy efficiency and energy management work together to reduce the energy demand.

1.5 Wastewater Treatment Technologies and Energy Use

These are the most important technologies used in wastewater treatment:

1.5.1 Activated Sludge Process

The activated sludge process (ASP) has been used in North America since early 1900s as secondary and tertiary biological treatments in WWTPs. There are many modifications of the activated sludge process but all consist essentially of an aerated biological reactor followed by a secondary clarifier. A generic activated sludge process configuration is presented in Figure 1-1.

Figure 1-1 Typical Conventional Activated Process Configuration

Conventional Activated Sludge Process

In conventional process, the aeration section consists in a channel. Settled primary effluent and returned activated sludge (RAS) enter the aeration section, and are mixed by diffused-air or mechanical aeration. During aeration period, adsorption, flocculation, and oxidation of organic matter occur. The suspended biomass also known as 'activated sludge' is subsequently separated from the wastewater in a secondary clarifier. Thickened biomass from the clarifier underflow is recycled to the aeration tank to maintain a desired biomass concentration. Diffused air used for achieving the required aeration and mechanical systems used for obtaining necessary mixing result in significant energy use.

Complete-mix Reactor

The aeration basin is a complete mix reactor (continuous-flow stirred tank reactor). The primary effluent and recycled activated sludge are introduced typically at several points in the aeration tank. The organic load on the aeration tank and the oxygen demand are uniform throughout the tank length.

Extended Aeration

In extended aeration (EA), there is no primary clarification. Extended aeration operates in the endogenous respiration phase of the microbial growth curve. This is the principal difference between

extended aeration and other activated sludge processes. Longer HRTs (18-36 hrs) provides greater process resilience to shock loading. The process is designed to operate at solids residence time (SRT) of 15 days or more. Extended aeration generates less solids from endogenous respiration and offers more overall process stability compared to other ASP. However, poor settling characteristic of effluent is a frequently encountered problem with extended aeration. This process is preferred for treatment of high strength wastewater. Extended aeration demands even more energy for aeration compared to conventional process.

Oxidation Ditches

The oxidation ditch consists of a ring- or oval-shaped channel and is equipped with mechanical aeration devices. Wastewater enters the ditch, is aerated, and circulates at about 0.8-1.2 ft/s. Oxidation ditches typically operate in an extended aeration mode with HRT > 10 hrs and SRT 10-50 days.

1.5.2 Sequencing Batch Reactors

The sequencing batch reactor (SBR) is a fill-and-draw non-steady state system using one or more complete mix reactors in which all steps of activated sludge process occur. According to the 1983 US EPA report, "the SBR is no more than an activated sludge system which operates in time rather than in space." There are two major classifications of SBRs: the intermittent flow (IF) or "true batch reactor," which employs all the steps in Figure 1-2, and the continuous flow (CF) system, which does not follow these steps. The IF-SBR accepts influent only at specified intervals. More than one operating strategy is possible during most of these steps.

Figure 1-2 SBR Sequence of Operation

Fill: Influent wastewater is added to the reactor and mixed with settled biomass from the previous cycle. Any or all of three variations: static fill, mixed fill, and aerated fill may be used depending on the operating strategy. An SBR serves as an equalization basin while filling, enables system to tolerate peak flows/loads. Because this system is closed to influent during the treatment cycle, two units may be operated in parallel.

React: Aeration turned on and fully aerobic conditions are maintained for a fixed period. Aeration may be provided by diffused air or mechanical devices.

Settle: Settle is typically provided under quiescent conditions. In some cases, gentle mixing during the initial stages may result in a clearer effluent and a more concentrated settled sludge.

Draw: The treated effluent wastewater is then discharged from the basin

Idle: Occurs between 'Draw' and 'Fill'. The step length varies depending on the flowrate and the operating strategy. Mixing to condition the biomass and sludge wasting can also be performed during this step.

In the CF-SBR, influent flows continuously during all phases. To reduce short-circuiting, a partition is normally added to separate the turbulent aeration zone from the quiescent area. CF systems hold the mass ratio of influent substrate to biomass constant by adjusting RAS flow rates continually as influent flow rates.

The performance of the SBR processes relies on the complex control system, which consists of a combination of level sensors, timers, and microprocessors. A higher level of sophistication requires complex tuning procedures and higher level of maintenance.

1.5.3 Energy Use in Activated Sludge Plants

According to 1994 survey, about 46% of the (total number of) wastewater treatment plants (surveyed) in Ontario were using conventional activated sludge process or variation of it, as a secondary treatment process, compared to 1% of the plants were using fixed film processes (Evans and Laughton, 1994). In terms of power usage, activated sludge plants were using 86% of the total energy consumption of WWTPs compared to 0.9% by fixed film processes. In terms of unit power usage, activated sludge processes were consuming between 0.33 and 1.13 KWH/m³ compared to 0.20 KWH/m³ by the fixed film processes. These show fixed film processes are more energy efficient than activated sludge treatment process.

Aeration is the most energy consuming process at WWTPs followed by pumping and these are the major components at activated sludge treatment plants. Aeration is the single largest energy consumer for activated sludge treatment facility, accounting for 54-97% of the total electrical energy requirements depending on the configuration of the treatment plant (Young and Koopman, 1991). In addition to that, the activated sludge process depends on sludge recycling for satisfactory operation. Return Activated Sludge pumping also consumes substantial amounts of energy at the activated sludge WWTPs, accounting for 6-15% of total energy consumption (Young and Koopman, 1991).

Elimination of clarifiers, clarification equipment, and RAS pumps, and use of sophisticated instrumentation to control aeration system could achieve some energy saving but increase capital and labor costs in SBR plants compared to activated sludge plants. Helmreich et al. (2000) reported 1.3-5.8 kWh/kg BODr of energy consumption by SBRs compared to 3.77-7.23 kWh/kg BODr reported by Young and Koopman (1991) for activated sludge processes.

1.5.4 Fluidized Bed Bioreactors

The fluidized bed bioreactor (FBBR) is an attached-growth biological technology, in which aqueous pollutants are digested and hydrolyzed within biofilm formed by bacteria on fluidized solid, also called bioparticles. The FBBR is similar to the packed-bed reactor in some respects, but the packing material is expanded by the upward movement of fluid (air or water) through the bed. The expanded porosity of the fluidized-bed packing material can be varied by controlling the flow rate of the fluid. The major advantages of fluidized bed bioreactors over other types of bioreactors such as activated sludge and trickling filters are

- A low washout rate of microbes from the system;
- A minimum sludge recycle and biomass recovery at a high substrate loading;
- No clogging of the system due to biomass accumulation; and
- A high and variable biomass concentration and large solid-liquid contacting area.

Figure 1-3 Schematics of Different Types of Fluidized Bed Bioreactor

Consequently, FBBRs yield high overall removal rates and low generation rates of excess sludge per unit reactor volume. Fluidized beds can be configured in a variety of forms in terms of their flow patterns and process characteristics, but with respect to wastewater treatment FBBRs are broadly classified as anaerobic and aerobic FBBRs. Schematics of FBBRs system are shown in Figure 1-3

Anaerobic FBBR

In anaerobic FBBR the wastewater is fed upward to a bed of 0.4 to 0.5 mm sand, lava rock, anthracite, activated carbon or similar kind of bioparticles. Bed depths vary depending on the organic loads. The specific surface area is about $1000 \text{ m}^2/\text{m}^3$ of reactor volume, which is greater than any of the other fixed film technologies. Upflow velocities are 30 to 36 m/h. Effluent recirculation is necessary to provide the fluid velocity within the necessary treatment residence times. Hydraulic retention times in FBBRs range from 5 to 20 min. As the biofilm increases in size, the packing changes density and moves to the center of the fluidized bed.

Aerobic FBBR

Aerobic FBBR can be classified in numerous ways. In this brief review we will comment on cocurrent externally and internally aerated configurations. In TFB, both pollutant removal and aeration occur simultaneously in the bioreactor. ETFBs, however, use a preaerator to enrich oxygen concentration in the feed wastewater, and a liquid-solid FBBR for substrate removal. In TFBs, bioparticles are supported by upward flows of gas and/or liquid. Aerobic FBBRs are frequently used to treat groundwater contaminated with hazardous substances. The unit energy consumption of FBBRs ranges from 1.25 to 2.5 KWH/kg BOD (Fan, 1989).

1.5.5 The Rotating Biological Contactors

The RBC Process is an attached growth secondary / tertiary biological wastewater treatment system, which consists of circular corrugated media mounted on a horizontal shaft. The media is slowly (1-10 rpm) rotated by mechanical or compressed air drive, while approximately 40% of the surface area is submerged in the wastewater.

Figure 1-4 Rotating Biological Contactor Treatment Process

Microorganisms present naturally in the wastewater adhere and multiply into a large microbial population on the rotating media surfaces. Wastewater flows through the reactor by simple

displacement and gravity. The RBC Process achieves high degrees of treatment for relatively short wastewater residence times. In rotation, the media carries the organisms, along with a film of wastewater, into the air where oxygen is absorbed. Rotation of the media results in alternating contact of microorganisms with wastewater and air and hence the dissolved oxygen and the soluble substrate material are uniformly available to the inner portion of the biomass through the mechanism of mixing as well as diffusion. Biomass continuously sloughs from the discs, becomes suspended solids and eventually sludge in a clarifier. The use of exposed and submerged stages in multiple tanks to create aerobic and anoxic conditions may be employed where nitrogen removal is required. In the smaller units multiple tanks are simulated in a single cradle using the partition also referred as baffles. When the RBC is used for nitrification and/or low BOD environments the high-density media may be used because of slower biomass growth and thinner biofilm. The RBC process has specific advantages over suspended growth processes. Its modular construction, low hydraulic head loss, and shallow tankage requirements allow it to be easily and economically integrated into existing facilities.

1.5.6 Low Energy Consumption - An RBC Advantage

The rotating biological contactor (RBC) process is similar to the activated sludge process in that it has some suspended culture in its mixed liquor and can produce high degrees of treatment and a sparkling clear effluent. However, the major difference is that more than 95% of the biological solids in this fixed-film system are attached to the media (Antonie, 1976), which eliminates the requirement of RAS pumping. In addition, the rotating biological contactor process utilizes oxygen from the air without the need for mechanical aeration mixing as required in the activated sludge process.

'Once-through' flow operation of RBC system with no sludge recycle and no mechanical aeration system, made them very simple and significantly energy efficient when compared to alternative treatment systems with sludge recycle. It is estimated that RBC systems would require 1/3–1/2 of the energy of an activated sludge system (Droste, 1997). Low energy requirements associated with RBC processes made them more attractive for treating domestic as well as industrial wastewaters.

1.6 Additional Aeration and Improved Performance

Historically wastewater treatment processes were used for the treatment of domestic wastes. Later when the usefulness of biological processes to treat industrial waste have been identified, these wastewater treatment processes were employed to treat high strength industrial wastewater as well as medium strength combined wastewater from domestic and industrial sources. When used for the treatment of high strength wastewater, with heavy organic loadings, initial stages of RBC can suffer problems such as low dissolved oxygen, heavy bio-film growth, the presence of nuisance organisms, such as beggiatoa, and a reduction in organic removal rates. (Chesner and Molof, 1977; Chesner and

Iannone, 1978; Hittlebaugh and Miller, 1981; USEPA, 1984; WEF, 1992; USEPA, 1993). Overloading, anaerobic growth, presence of beggiatoa and lack of operator control comprise the majority of problems and criticisms encountered at RBC facilities experiencing difficulties.

Overloading problems can be overcome by removing the baffle and enlarging first stage to reduce surface loading, and increase oxygen-transfer capability by increasing the rotational speed. Other approaches include supplemental air system, step-feeding, or recycling from the last stage (Tchobanoglous et al., 2003).

Increasing rotational speed is not an economic alternative. This seemingly small adjustment would double the power consumption, doubling the operating costs. Baffle adjustment for staging alternative does not easily lend itself to short term problems. This option entails draining tankage to remove baffles and would most likely be accomplished after the problem had passed. Enlargement of the first stage, step feeding and recycling, ends up with an additional RBC and major modifications of the existing system. In contrast, supplemental aeration is a relatively easy solution especially in the case of modification of the existing facilities.

Studies on RBCs using a supplemental air system have shown improved efficiency in terms of organic and $\text{NH}_3\text{-N}$ removal as well as stage-DO concentrations (Surampalli and Baumann, 1995; Surampalli and Baumann, 1997; Kargi and Uygur, 1997a). The improved oxygenation provides the operator one more level of control with the system - air flow rate. These can be adjusted as required during seasonal peaks or unanticipated discharges from heavy industrial users. While this would permit the operator during low periods to take advantage of additional power savings, the major thrust is the ability to safeguard against unforeseen heavy loads.

1.7 Objective and Scope of the Thesis

The main objective of this study was to optimize the RBC performance by minimizing energy consumption for unit oxygen demand removed. The secondary objective was to investigate the energy consumption per unit oxygen demand removal at different combinations of disc rotational speed and rate of aeration, and to optimize the operating parameters.

The use of additional aeration eliminates the DO limiting situation, improves the overall performance for simultaneous BOD removal and nitrification, and provides the operator operational flexibility. But on the other hand, the aeration system is energy inefficient compared to the aeration through media rotation. For better energy management, it is required to maximize the performance of the RBC system for unit energy used or minimize the energy used for unit mass of oxygen demand removed. This can be achieved by balancing the supply of air through media rotation and from supplemental aeration system, and minimizing energy consumption per unit oxygen demand removed.

Scope of this study:

1. Determine the energy consumption by a commercial RBC of 3000 L/day capacity operating with tap water without biofilm.
2. Determine steady state dissolved oxygen profiles in a modified rotating biological contactor without biofilm at different combinations of angular velocity and rate of aeration.
3. Assess the energy utilization per unit oxygen demand removal (unit energy consumption) at different combinations of angular velocity and rate of aeration, using synthetic wastewater of fixed chemical characteristics and flow rates.
4. Develop a mathematical model to determine energy utilization with fixed influent characteristics and flow rates.
5. Determine local optima if any for unit energy consumption, with fixed influent characteristics and flow rates.

2. LITERATURE REVIEW

2.1 Brief History

Fixed-biofilm reactors are widely used as treatment systems for liquid organic wastes originating from industrial as well as domestic sources. The application of biofilm technology in wastewater treatment originated from the first full-scale operation of low rate trickling filters in early 1880s in Wales (Lazarova and Manem, 2000). The rotating biological contactor goes back to 1900, when German scientist, Weigand first envisaged the contactor for wastewater treatment. His patent for the contactor (Weigand, 1900) describes a cylinder consisting of wooden slats. Nothing much happened until Imhoff (1926) and Bach (1934) tested them, as substitutes for the Emscher filter. These contactors experienced severe problems with clogging of the slats and were not investigated further.

During late 1920s, U.S. scientist Maltby talked about the innovation of the 'biological wheel' which consisted of a series of rotating paddle wheels (Allen, 1929; Winkler, 1981). In the same year, Doman (1929) reported on his testing of rotating metal discs. This was the first time that discs were investigated as the contact media, but the results were not encouraging, and no further work was done in the U.S. In Europe, nothing further developed until the late 1950s. In that decade, investigators at Stuttgart University, West Germany (Hartman, 1960; Popel, 1964), attempted to improve the secondary treatment process, by experimenting with wooden and plastic flat discs. The process development work of Hartman and Popel, together with expanded polystyrene, a new material of construction at that time, resulted in a new commercial wastewater treatment process. These discs were about 1 cm thick and were placed 3 cm apart. This arrangement could provide a surface area of only about 25 m²/m³ of reactor. At 50% submergence, the surface area would only be 50 m²/m³ of liquid.

In 1957, J. Conrad Stengelin began to manufacture 2 and 3 meter diameter expanded polystyrene discs in West Germany. The first commercial installation went on stream there in 1960, and use of the process spread quickly throughout Europe because of its simplicity and low power consumption. Although offering greater simplicity in operation and economical on energy consumption, the process was restricted to relatively small installations.

After 1960, further development of the rotating biological contactor stopped in Europe, but between 1960 and 1965 the U.S. company, 'Allis-Chalmers' did much development of rotating discs, without knowledge of the previous work. Later they reached a licensing agreement with the German manufacturer for manufacturing and sales distribution under the trade name BIO-DISC in both the U.S. and Europe. The first commercial installation in U.S. went into operation at a small cheese factory in 1969. In 1970, Allis-Chalmers sold its RBC technology to the Autotrol Corp. At that time, the polystyrene discs were still not competitive with the activated sludge process. Use of the expanded-polystyrene disc construction continued with some commercial success, but, as in Europe, it was limited to small domestic and industrial waste applications.

Then, in 1972, came an important breakthrough, the development of a more compact disc constructed from corrugated sheets of polyethylene, one with much more surface area for a given volume. This increased the surface area density from $16 \text{ ft}^2/\text{ft}^3$ ($52.5 \text{ m}^2/\text{m}^3$) for the polystyrene disc to $37 \text{ ft}^2/\text{ft}^3$ ($121.4 \text{ m}^2/\text{m}^3$) for the new corrugated media, provided as much as 100,000 ft^2 of surface area available on one contactor. Because the media was corrugated and the discs were closely packed, the identity of the discs was lost. During the initial research work at Allis-Chalmers, the process was referred to as the 'Rotating Biological Contactor' (RBC) and later it is also referred with the same name in the literature. Since this name is not associated with any particular manufacturer, it became a generic term.

In mid 1970s, there were nearly 100 installations in U.S. and Canada. Fifteen had more than 1 mgd capacity. Now there are numerous RBC units operating world wide (ASCE Manual, 1977; Lumbers, 1983; Tchobanoglous et al., 2003; Findlay, 1993; Mba et al., 1999), ranging from small units serving individual dwellings to very large ones having the capacity to treat flows of several million liters per day. The early history of RBCs reliability particularly in the pre-1980 period was marked by an exceptionally high rate of shaft and several other mechanical failures world wide (Brenner et al., 1984; Tchobanoglous et al., 2003; McGhee, 1991; Horan, 1991), due to poor knowledge of loading factors, minimal understanding of the fatigue problem including use of unconservative design stress allowables, bad fabrication practices, improper preventive maintenance and microbial corrosion (Mba et al., 1999). In fact, one prominent manufacturer experienced an almost 5% failure rate over a 5 year period on 3000 RBC installation in municipal waste treatment facilities. (Bernard Ross et al., 1994). More recent designs have eliminated most of these problems (Griffin and Findlay, 2000).

Many modifications of the RBCs have been introduced till date, AQUACAN[®] series is one of them. AQUACAN[®] was developed in 2002 by Ecotechnos Inc. Ecotechnos is one of the few Canadian companies engaged in designs, manufacture and service of small and medium-sized advanced wastewater treatment units for high strength industrial and domestic wastewater. AQUACAN[®] series are very powerful, compact, easy to maintain, modular small treatment units generically known as 'Active Biological Contactors (ABC)'. ABC combines the advantage of RBC and the mass transfer

effects of three-phase fluidized beds near the air distributors in one compact design. These provide the operator with one more parameter to control and hence relatively more flexibility to treat wide ranges of flow and characteristics of wastewater. In 2003, this technology was successfully tested with high strength winery wastewater (Alvarez-Cuenca M., 2004). In 2004, AQUACAN® technology was certified under License # ETV2004-43, by the Canadian government under 'Environmental Technologies Verification' (ETV) program. In June 2005, this technology was assessed and approved by Standard Development Branch, Ontario Ministry of Environment. The AQUACAN® technology is patented in both Canada and the United State. (Alvarez-Cuenca M., 2006)

2.2 The RBC Technology

The rotating biological contactor is a fixed film bioreactor designed for secondary and tertiary treatments of domestic as well as industrial wastewater. Table 2-1 lists some typical applications of RBC.

RBCs consist of a series of closely spaced circular discs mounted on a longitudinal shaft with at least 40% of the discs submerged in the wastewater. When the process is placed in operation, the microbes naturally present in the wastewater begin to adhere to the rotating surfaces and grow there until the entire surface area of the discs is covered with a 1 to 4 mm layer of biological slimes, referred to as a fixed-film. RBCs employ the attached film of microorganisms also known as biofilm, to absorb and metabolize the organic matter in wastewater. The set of discs on which the biomass grows is called the media. Now most of the manufacturers offer media with a lattice structure and are typically made of light weight polymeric material such as PVC, polystyrene, or polyethylene. This more complex structure offers more surface area per unit volume. High-density polyethylene (HDPE) has been the standard for plastic media since the early 1970s (WEF and ASCE, 1998) and is most commonly used.

As indicated above during the periodic submersion, the attached microbial mass is brought into contact with wastewater being treated for absorption of the organic pollutants and nutrients. While exposed to the atmospheric air, a film of wastewater is carried along the biofilm, which trickles down the surfaces and entrains oxygen from the air by diffusion. As the discs complete their rotation, this film mixes with the reservoir of wastewater, adding to the oxygen in the reservoir and mixing the treated and partially treated wastewater (Suga and Boongorsrang, 1984). Unused dissolved oxygen in the wastewater film is mixed with the contents of the mixed liquor, which maintains a mixed-liquor dissolved-oxygen concentration. Thus, in principle, these contactors enable contact between the gas phase and the liquid phase. Hence, in this process, the microorganisms are provided with a steady supply of both organic material as well as oxygen for their cell growth and respiration. Cell growth and substrate oxidation are generally considered to be coupled reactions, i.e. substrate removal occurs because of cell growth (Grady Jr., 1983).

Table 2-1 Applications of RBC

Applications	References
Municipal/Domestic Wastewater Treatment	Hiras et al. (2004); Tawfik et al. (2002); Castillo et al. (1997); d'Antonio et al. (1997); Nue (1992)*, Klees et al. (1992)*
Treatment of Saline Wastewater	Kargi and Uygur (1997b); Kargi (2002)
Groundwater Treatment	Mohseni-Bandpi et al. (1999); Mohseni-Bandpi and Elliott (1998) (1996); Engelder et al. (1990) $\square$
Treatment of Landfill Leachate	Mckim et al. (1993)*, Forgie (1988)*, Peddie et al.(1985)*
Treatment of Wine and Food Based Processing Wastewater	Alvarez-Cuenca M. (2004); Liu et al.(2004); Malandra et al. (2003); Wahaab and El-awady(1999); Laughton(1990)*; Pajak and Loehr(1976)
Cheese Processing Wastewater Treatment	Hammer (1982)*
Shellfish Wastewater Treatment	Hudson et al. (1976)*
Cannery Wastewater Treatment	Max et al. (1970)*
Salmon Canning Wastewater Treatment	Clagget (1973)*
Paper Industries Wastewater Treatment	Bryant et al. (1997)
Dyestuff Wastewater Treatment	Karapinar and Kargi (2002); Ge et al. (2004); Van Driessel et al. (2001); Abraham et al. (2003)
Waste Gas Treatment	Vinage and Rudolf von Rohr (2003a) (2003b)
Metal & Heavy Metal Removal	Costley and Wallis (2001a) (2001b); Stott et.al. (2001); Patil and Paknikar (2000) ; Ganguli and Tripathi (2002); Russell et al. (1986)*
Biodegradation of Hydrocarbons	Brar and Gupta (2000); Zheng and Obbard (2002); Yamaguchi et al. (1999); Hsueh et al. (1991)*; Huang, et al. (1985).*
Biodegradation of Phenol	Alemzadeh et al. (2002); Israni et al. (2002)
Biodegradation of Chlorinated Phenol	Alleman et al. (1995); Swaminathan and Ramanujam (1999); Radwan and Ramanujam (1997a); Zouari et al. (2002); YucelToruz(1988)*
Biodegradation of Toluene	Alemzadeh and Vossoughi (2001)
Petroleum Refineries Wastewater Treatment	Fang et al. (1993) $\square$
Fermentation	Saha and Takahashi (1997); Wang, (2000)
For Aquaculture System	Lee et al. (2000); delos Reyes Jr. and Lawson (1996)
Anaerobic Treatment of High-strength Wastewater	Lu et al. (1995)

* (Adopted from: Alvarez-Cuenca et al. (1993) - see this paper for details of references)

$\square$ (Adopted from: Castilla Rodriguez (2001) - see this thesis for details of references)

Shear forces are exerted on the biofilm due to the rotation of the discs through the wastewater. These shear forces strip the discs of excess biomass, and provide a sufficient degree of mixing to keep the sloughed biomass in suspension until it is carried out of the reactor tank with the bulk flow, thus eliminating the possibility of reactor clogging due to biomass accumulation.

Operating in this manner, the discs serve several purposes, such as:

- To provide media for buildup of attached microbial growth;
- To bring the growth into contact with the wastewater;
- To aerate the wastewater and suspended microbial growth efficiently, in the wastewater stages;
- To provide a positive means of continuously stripping excess biomass; and
- To thoroughly mix each stage of treatment and keep sloughed solids in suspension.
- To enhance the flow of the wastewater to the next stage.

Rotation of the media also provides turbulence at the interface between biomass and wastewater, so that dissolved oxygen and soluble substrate are available to the inner portion of the biomass through the mechanism of mixing as well as that of diffusion.

RBCs are often designed as a series of stages. Each stage of RBC simulates a completely mixed reactor and the residence time distribution through multiple stages approaches plug-flow. In each stage, the rate of biological growth and the rate of stripping excess biomass eventually reach at a dynamic equilibrium. The attached biomass contains approximately 50,000 to 100,000 mg/l suspended solids. Stripped biomass with wastewater in the stage, results in mixed-liquor suspended solids concentration 10,000 to 20,000 mg/l (Antonie, 1976). This large inventory of fixed microbial population on media permits high degrees of treatment to be achieved for relatively short wastewater residence times. Thus the rotating biological contactor (RBC) represents a viable means for the secondary treatment of both municipal and industrial wastewaters (Banerjee, 1997a) by exploiting the advantages of both fixed film and suspended growth systems. The growth that is stripped off is in the form of relatively large aggregates or flocs dense biomass, which settle rapidly in a final clarifier.

2.3 Advantages and Disadvantages of RBCs

The following are the advantages and disadvantages of RBCs used in the simultaneous organic carbon and nitrogen removal of high strength wastewater:

2.3.1 Advantages

- Series of stages in RBCs makes them suitable for simultaneous removal of organic carbon and nitrification without intermediate clarification and sludge recycle.
- Modular configuration makes their installation and space requirements more economic
- No sludge or effluent recycle streams.
- Recover faster than activated sludge from toxic shocks.
- RBCs can provide a very high degree of treatment. The addition of an aeration system to the RBC processes, allows much higher level of the treatment, and more flexibility.
- RBCs can handle organic overloads well because of large biomass inventory in the system. Attached growth in RBCs makes them much less likely to suffer from biomass washout under adverse conditions.
- No short-circuiting in the biological reactor. No bulking, no foaming, or no floating of sludge to interfere with a plant's overall efficiency.
- Sloughed biomass settles well and can be more reliably removed than the solids from an activated sludge process. Good settling characteristics of sloughed biomass, makes design and operation of clarifier far less critical than in other treatment processes.
- Because most RBCs operate with a net increase in hydraulic head, pumping which would otherwise have been needed may be obviated.
- High oxygen transfer efficiency of RBCs, makes them consuming up to 50 % less energy than activated sludge process for the same level of treatment.
- RBCs require less area than most other comparable processes. In laying out a plant, RBCs offer advantages beyond their relatively low area requirements. Fewer excavation requirements for RBCs make them suitable in high water table areas.
- RBCs are simple to operate and require less skilled labor. Mechanical simplicity of RBCs and the low speed of the units reduce their maintenance requirements and prolong their lives.
- The rotating disc process lends itself well to upgrading existing treatment facilities.
- Finally, RBCs are versatile both in the functions which they perform and in the flexibility with which they can be configured.

2.3.2 Disadvantages

RBC technology is not without its share of problems.

- RBC does not exhibit the economies of scale observed with other non-modular technologies. Although low maintenance costs and area requirements are often cited as RBCs advantages,

they are strictly proportional to plant capacity, exhibiting none of the economies of scale observed with other non-modular technologies.

- Similarly, area requirements for RBC installations are proportional to plant capacity so that an advantage for small and medium-scale plants becomes a liability in very large capacity applications.
- Anaerobic and limited DO conditions can result in the growth of sulphide oxidising white biomass composed of Thiotrix or Beggiatoa. Overloading of the first stage of a RBC installation can cause odors and inadequate removals. Extra surface area can be provided in the first stage to alleviate the overload conditions. When the overloads are episodic, equalization upstream from the RBC reactor can be as useful with this technology as with others.
- Operational experience is shorter than that found in activated sludge.

2.4 Carbonaceous Material Removal in RBC

The purpose of biological treatment relative to carbonaceous material is the conversion of soluble organics to particulate bacteria, and stable end products. When organic waste is treated by an RBC, the attached biomass carries out the conversion in general accordance with the symbolic stoichiometry shown in equations (2-1) and (2-2)

Oxidation and Synthesis:

Endogenous Respiration:

COHNS symbolizes the organic matter in wastewater. Although the endogenous respiration reaction results in relatively simple end products and energy, stable organic end products are also formed. From equation (2-2), it can be calculated that, if all of the cells ($C_5H_7NO_2$) can be oxidized completely, the ultimate BOD of the cells is equal to 1.42 times the concentration of cells.

2.5 Nitrification in RBC

Ground and surface water pollution are grave problems facing many developing countries due to discharge of excess organics, nitrogen and phosphorous substances. Nitrogen is present in the majority of wastewater streams including municipal and industrial wastewaters. Proteins and amino acids the main sources of nitrogen in municipal wastewater eventually break-down to ammonia. Ammonia is toxic to aquatic organisms, especially the higher forms such as fish even at concentrations as low as 0.5 mg/l (Barnes and Bliss, 1983).

Nitrification is used to control the ammonia level in the wastewater effluent. The principle of biologically induced nitrogen removal is wholly based on the activity of populations of autotrophic nitrifying bacteria and their capability to sequentially oxidize nitrogen from ammonia to nitrate. The two microbial genera usually associated with nitrification are *Nitrosomonas* and *Nitrobacter*. Both genera of organisms are autotrophic nitrifying bacteria indicating that energy for growth is derived from the oxidation of inorganic nitrogen. The oxidation of ammonia to nitrate is a two step process requiring both organisms for the conversion. First ammonium ion is oxidized into nitrite by bacteria commonly generalized to be of genus *Nitrosomonas*:

The acid produced is neutralized by a reduction of the alkalinity, which is measured as bicarbonate equivalents:

Bacteria that have previously been generalized to be of genus *Nitrobacter* further oxidize the nitrite into nitrate:

The overall oxidation of ammonia by these organisms is given by the following equation (2-6) :

As ammonia is oxidized to nitrate, carbonate is utilized and converted to carbonic acid. This microbiologically induced change results in the destruction of alkalinity at a rate of 7.1 mg (as CaCO_3) per mg of ammonia oxidized (USEPA, 1975a). As the nitrification process proceeds, the pH of the wastewater may drop as low as pH 6.0, and adversely impact the rate of nitrification. Optimum pH for effective nitrification usually falls between 7.0 and 9.0. (Sawyer et al., 1973; Engel and Alexander, 1958), and nitrification ceases at or below pH 6.3 to 6.7 (Painter, 1970). Poduska and Andrews (1974) have shown that abrupt changes in pH from 7.2 to 5.8 markedly reduced the ammonia oxidation by nitrifiers while the reversal in pH restored the original nitrification rate. This decrease in pH can be minimized by aeration to strip CO_2 from the wastewater, or by ensuring the presence of excess alkalinity.

Besides the oxidation and generation of energy from inorganic nitrogen, some of the ammonium ions are assimilated into cell tissue. The biomass synthesis reaction can be represented as follows:

The overall oxidation and synthesis reaction can be represented as follows:

The oxygen required to oxidize ammonia to nitrate in overall oxidation and synthesis equation (2-8) is 4.3 mg O_2 / mg of ammonium nitrogen, which is in close agreement with the value of 4.57, which is often recommended for design calculations. The value of 4.57 is derived from equation (2-6) in which cell synthesis is not considered. (US EPA, 1975a)

The optimum temperature for nitrification is 30°C with a range of $28\text{--}35^\circ\text{C}$. (Buswell et al., 1954; Laudelout and vanTichelen, 1960; Balakrishnan and Eckenfelder, 1969; Haug and McCarty, 1971; Haug and Hopson, 1974). Temperature influences heterotrophic and autotrophic microorganisms, thereby affecting organic removal and nitrification efficiencies. The nitrification rate is more temperature sensitive than the rates for organic carbon removal (Busch, 1971). Nitrification rates decrease about 50 % for each 10°C drop in wastewater temperature below 30°C (USEPA, 1975a).

The organic carbon removal efficiency is less likely to be affected by temperature changes, probably due to microbial population diversity and other system constraints. Organic carbon removal rates for

fixed-film processes decrease about 25% for each 10°C drop in wastewater temperature below 30°C. However, the actual temperature effect on a biological process is probably characteristic only of that system.

In addition to nitrification, microorganisms other than the nitrifiers require nitrogen for growth. The amount of nitrogen assimilated during oxidation of carbonaceous material has been generally placed at 5% of the oxygen demand (i.e., BOD:N = 20:1). The consequences are two fold: (1) nitrogen must be present for biological oxidation of carbonaceous material, and (2) removal of ammonia-nitrogen during biological treatment of wastewaters may be due to assimilation, not necessarily due to nitrification.

2.6 Nature of the Biofilm in RBC

The initial phase in biofilm development involves the adsorption of organic compounds over the material which will be colonized (Trulear and Characklis, 1982). This initial organic layer is a prerequisite for the later microbial attachment (Baier, 1972; Fletcher, 1980). The biofilm development begins after that phase. The biofilm is visible a few days after the start-up of the reactor (Bishop and Kinner, 1986). The duration of this adherence phase will depend on several factors: the nature of the support (Zobell, 1943; Wood, 1950; Fletcher and Floodgate, 1976; Beachy, 1981), the surface charge, the nature and the concentration of the feed, etc. The initial surface colonization occurs at cavities on the support, which has a surface roughness favorable for this development.

During the growth phase, a quick biofilm development occurs due to the growth of microcolonies and the adherence of new bacteria (Capdeville and Nguyen, 1990), so that at the end of this phase, the surface is totally covered by the biofilm with a complex structure of microbial cell clusters (de Beer et al., 1994). This growth phase can be divided into two steps, the first one with a logarithmic biofilm growth and the other one with a constant biofilm accumulation rate, which continues until its partial detachment and the steady-state biofilm thickness is reached.

Figure 2-1 Nature of Biofilms in Rotating biological Contactor and Trickling Filter
(adopted from Antonie, 1976)

The nature of the attached biological growth in RBCs is very different from that developed in trickling filter systems. Figure 2-1 is an attempt to depict this difference. In trickling filters, the growth is uniform in thickness and often has a gelatinous appearance, which is why the growth is usually referred to as slime. In contrast, the biomass of the rotating contactor is shaggy with many macroscopic filaments, which project outward into the adjacent film of wastewater. This provides an active biological surface area much larger than just the surface area of the support medium. It also enables the substrate and dissolved oxygen to reach a greater portion of the biomass and render it aerobically active. The shaggy growth is not due to the presence of any species of organisms but to the action of rotation. The continual drag from being rotated through the wastewater, and the draining of entrained wastewater when rotated up into the air, cause the growth to form the elongated macroscopic filaments. The shaggy growth increases in length and thickness until it can no longer withstand the shear exerted by rotation. The growth that is stripped off is in the form of relatively large aggregates of dense biomass, which settle rapidly in a final clarifier.

The function and structure of the biofilm depend on the conditions it is exposed to. This shaggy structure is most apparent in the initial stages of the media, where the BOD concentrations are higher and the growth is thicker. The biofilm thickness gradually decreases in later stages, where BOD concentrations are lower and growth is slower. The initial stages of the media receive the highest concentration of organic matter and develop cultures of filamentous and nonfilamentous bacteria. As the concentration of organic matter decreases in subsequent stages, higher life forms including nitrifying bacteria begin to appear, along with various types of protozoans, rotifers, and other predators.

In heterotrophic biofilms, facultative bacteria are the predominating organisms, and they decompose organic material in the wastewater along with aerobic bacteria and anaerobic bacteria in oxygen depleted interior of the biofilm. If the organic loading is high, the aeration is sufficient and the hydraulic loading is moderate the biofilms may become very thick (>10 mm). Because of a relatively slow growth of nitrifiers, autotrophic nitrifying biofilms are generally thinner (<1 mm) than heterotrophic biofilms. An increased thickness of biofilm does not necessarily give higher substrate removal rates. The substrate removal rate depends on the active layer of biomass. The depth of the active layer is controlled by the diffusivity of dissolved oxygen into the biofilm. (Venkataraman and Ramanujam, 1998) In places where the biofilm is thick and fairly homogenous, the oxygen used by the aerobic bacteria only penetrates the outer 0.1– 0.3 mm. Bungay (1969) found that the respiration ceased at a depth of 50-150 μm , depending upon the substrate concentration in the wastewater. Underneath, the environment is anaerobic and if soluble biodegradable organic matter and nitrate is present, denitrification of nitrate to nitrogen may occur (Dalsgaard and Revenspech, 1992; Wik, 1999).

2.7 Oxygen Mass-Transfer in RBCs

The ability of an RBC for oxygen transfer to and within the biofilm affects the overall performance of the process. (Bintanja et al., 1975; Zeevalkink et al., 1979; Shieh Wen, 1982; Matsuo et al., 1985) The mass transfer aspects have been studied for a set of plain discs by several researchers and different models have been proposed. The system was first studied without biomass, where only physical transfer occurs. Various authors have provided experimental values for the physical transfer (Bintanja et al., 1975; Zeevalkink et al., 1979; Kim and Molof, 1982). Several mathematical models have been suggested to quantify this transfer.

Yamane and Yoshida (1972) were among the first researchers to theoretically analyze mass-transfer aspects. They modeled the oxygen transfer to the liquid film with varying thickness δ , on the disc surface. Yamane and Yoshida (1972) and Bintanja et al. (1975) assumed that the liquid film entrained on the disc surface is stripped off and completely mixed with the bulk liquid as soon as the disc re-enters the liquid. Ravetkar and Kale (1981) assumed a uniform film thickness on the disc surface. In addition, they considered that the disc surface wetted by the liquid depends on the submergence of the disc. They correlated experimental data on the mass-transfer coefficient using the following equation (2-9).

$$\frac{K_L D}{D_m} = 1.59 \left(\frac{D^2 N}{D_m} \right)^{0.5} \quad (2-9)$$

Where,

K_L	Gas-Liquid Mass-transfer coefficient for the liquid film exposed to air, m/s
D	Diameter of the disc, m
D_m	Molecular diffusivity of oxygen, m^2/s
N	Speed of rotation, s^{-1}

Castilla Rodriguez (2001) investigated oxygen mass transfer in the RBC without any biomass. He used the same RBC used in this research. He reported the following empirical equation (2-10) correlating oxygen mass transfer coefficient (K_L) and the angular velocity of the discs (ω).

$$K_L = (2.49\omega + 1.45)10^{-6} \quad (2-10)$$

Zeevalkink et al. (1979) suggested another mechanism of transfer enhancement related to increased oxygen diffusion in the liquid film induced by the oxygen consumption in the biofilm. Be that as it may, the existence of this enhancement factor was experimentally demonstrated by Kim and Molof (1982), later by Paolini (1986) and, by Radwan and Ramanujam (1995).

Kim and Molof (1982) suggested oxygen transfer in RBC systems occur based on the following mechanisms:

- Oxygen absorption at the liquid film flowing over the disc's surface during the air exposure cycle,
- Direct oxygen transfer at the air-disc liquid interface, with this diffusion being the result of the mixing effect created by the rotating discs,
- Direct oxygen absorption by the microorganisms during the air exposure of the discs.

The authors were the first to focus on the possible increase in transfer coefficient resulting from biological activity. They conducted experimental measurements of the overall mass-transfer coefficient and correlated the volumetric mass-transfer coefficient using the following equation (2-11).

$$K_L a = 0.0011 \left(\frac{\omega^{1.5} D^{0.5}}{d} \right)^{0.732} \quad (2-11)$$

Where,

$K_L a$	Gas-Liquid volumetric oxygen mass-transfer coefficient, min^{-1}
ω	Angular speed of rotation, min^{-1}
D	Diameter of the disc, inches
d	Disc spacing, inches

The term in parenthesis was defined as the volume renewal number. This is not a dimensionless number. The units of ω should be revolutions per minute, whereas those of disc diameter, D and contiguous disc spacing d should be inches; thus, the calculated $K_L a$ has units of minutes inverse.

Paolini (1986) investigated the effect of biomass on the oxygen-transfer rate in RBC systems, in terms of dimensionless enhancement factor, E .

$$E = \frac{K'_L a V_L \left(1 + \frac{A_e}{A_i}\right)}{K_L a_w V'_L} \quad (2-12)$$

Where,

E	Enhancement of the oxygen transfer rate caused by the biomass
$K'_L a$	Overall biochemical oxygen mass-transfer coefficient in the reactor, min^{-1}
$K_L a_w$	Overall physical oxygen mass-transfer coefficient in wastewater ($=\alpha K'_L a$), min^{-1}
A_e	Exposed area of the discs, m^2
A_i	Submerged area of the discs, m^2
V_L	Effective volume of liquid in the reactor, m^3
V'_L	Effective volume of liquid in reactor without biomass, m^3

It was observed by Paolini that, in the presence of microorganisms, there is a very large enhancement in the rate of mass transfer. Paolini found that the enhancement factor was in the range of 2 to 10. Radwan and Ramanujam (1995) obtained an enhancement factor in the range of 2.5 to 4. The high enhancement factor confirms the large effect of the biological activity on oxygen transfer and the efficiency of rotating biological contactors as aerators.

Boumansour and Vassel (1998) carried out detailed investigations on the mass-transfer rate and its enhancement due to the presence of the biofilm. The mass-transfer coefficient was calculated by measuring the dissolved oxygen concentration in the liquid phase. They showed that their experimental data as well as the data of previous workers could be correlated using the following expression, and hence, their correlation would be expected to predict mass-transfer coefficients more accurately.

$$\frac{K_L D}{D_m} = 2.673 (ND^2 \rho)^{0.769} \left(\frac{N^2 D}{g} \right)^{0.135} \left(\frac{H}{r} \right)^{0.865} \quad (2-13)$$

Where,

K_L	Gas-Liquid Mass-transfer coefficient for the liquid film exposed to air, m/s
D	Diameter of the disc, m
D_m	Molecular diffusivity of oxygen, m^2/s
N	Speed of rotation, s^{-1}
g	Acceleration due to gravity, m/s^2
H	Height of submergence of the disc from the bottom, m
r	Radius of the disc, m
ρ	Density of wastewater, kg/m^3

This correlation predicts, $K_L \propto N^{1.039} D^{0.673}$ at constant H/r ratio.

2.8 Concentration Profile in the Series of Continuous Stirred Tank Reactors

In the case of smaller flow rates and longer residence times, each stage of the RBC reactor can be easily assumed as individual continuous stirred tank reactors (CSTR). Equation (2-15) is a general form of the mass-balance equation for a CSTR.

$$\text{Accumulation} = \text{inflow} - \text{outflow} + \text{generation} \quad (2-14)$$

$$\frac{dC}{dt} V = QC_0 - QC + r_c V \quad (2-15)$$

Where,

Q	Volumetric flow rate, m^3/s
C	Concentration of a reactant in the reactor at time t , g/m^3
C_0	Initial concentration of a reactant, g/m^3

V	Reactor volume, m^3
r_c	Rate of reaction, g/m^3s

Under steady state conditions, the rate of accumulation term is equal to zero ($dc/dt = 0$) and by assuming first-order removal kinetics ($r_c = -kC$) equation (2-15) can be simplified as given below.

$$C = \frac{C_0}{\left[1 + k\left(\frac{V}{Q}\right)\right]} = \frac{C_0}{[1 + k\tau]} \quad (2-16)$$

Where,

k	First-order reaction rate constant, s^{-1}
τ	Theoretical Residence Time (V/Q), s

If the reaction rate, k is known this equation can be used to predict the concentration in each stage of the RBC. Equation (2-17) can be used in case for n stages of equal volume to predict the concentration after n stages.

$$C_n = \frac{C_0}{\left[1 + \left(\frac{kV}{nQ}\right)\right]^n} = \frac{C_0}{\left[1 + \left(\frac{k\tau}{n}\right)\right]^n} \quad (2-17)$$

2.9 Simultaneous Organic Removal and Nitrification in RBC

Historically, rotating biological contactors have been used to remove organic carbon pollutants from wastewater, but in the last twenty years this has been expanded to include nitrification and denitrification of wastewater. Once through rotating disc process appear to be especially suited to obtaining nitrification. Because the process is constructed in a series of stages, the fixed cultures that develop in each successive stage become adapted to treating the wastewater as it undergoes a progressively increasing degree of treatment. This provides a valuable feature for achieving simultaneous organic removal and biological nitrification. Table 2-2 lists various studies carried out for simultaneous carbon removal and nitrification.

Table 2-2 Selected Cases of Simultaneous Carbon and Nitrogen Removal and/or High Strength Wastewater Treatment with RBC

References	Type of Wastewater	Loading					HRT
		Hydraulic	Organic		NH ₄ ⁺ -N		
			X 10 ⁻³ m ³ /m ² .day	mgCOD/L	gCOD/m ² .d	mg/L	
Gupta and Gupta, (1999)	Synthetic Domestic	8.88-26.66	259.8	2.31-6.93	26.18	0.23-0.70	Hrs
Gupta and Gupta (2001)	Synthetic High Strength Domestic Wastewater	6.66-10.66	500-1000	10-32	54.4-110.8	1-3.35	6-18
Hiras et.al. (2004)	Synthetic Municipal	27-68	294-822 (COD) 199-521 (BOD)	3.4-18 (COD) 3.8-10 (BOD)	27-73	0.24-1.8	3.6-9
Radwan and Ramanujam (1997b)	Synthetic High Strength	30	1000-1557	29.99-46.7	20-185	0.66-5.5	4.3
Watanabe et.al., (1995)	Synthetic	68.115	2-6 (C:N Ratio)	-	25	0.7	5.5

The heterotrophic bacteria responsible for carbonaceous BOD removal have an energetical advantage over the autotrophic nitrifying bacteria. This means, in an environment where oxygen and both ammonium and easily biodegradable organic matter are available in high concentrations, the nitrifiers will have a lower growth rate than the heterotrophs. Since they compete for oxygen as well as space the nitrifiers loose the competition and will eventually be out-spaced. This has been observed in a number of studies over the last few decades (Kissel et al., 1984; Wanner and Gujer, 1984, 1986; Furumai and Rittmann, 1992; Rittmann and Manem, 1992; Okabe et al., 1995; Wik and Breitholtz 1996).

The development and predominance of ammonia-oxidizing organisms in the process has been found to be primarily a function of BOD concentration. Nitrification usually does not begin until the soluble-BOD₅ level and corresponding large populations of heterotrophic organisms have been adequately lowered. As carbonaceous matter is removed by biological treatment, and as the BOD concentration approaches an undetermined BOD concentration, nitrifying organisms begin to establish themselves in the process. Ammonia oxidation begins and increases rapidly as the BOD concentration is further reduced. The maximum nitrification rate occurs when soluble BOD drops to 10 mg/l (WEF & ASCE 1998). The maximum nitrification rate strongly depends on dissolved oxygen in the bulk fluid, provided that the ammonium concentration exceeds 3–5 mg/l. Thus, the majority of the carbonaceous BOD is satisfied in the initial stages of discs. This allows nitrifying bacteria to predominate in subsequent stages for the achievement of high degree of ammonia oxidation. This occurs without intermediate clarification or sludge recycle. High degrees of nitrification can be achieved in as few as two stages of treatment. However, to maximize BOD and suspended solids removals while obtaining nitrification, it is more efficient to operate in four stages of treatment. (Antonie, 1976).

Figure 2-2 Basic Phenomena in the Biofilm

In RBC, nitrifying organisms may be interspersed among the heterotropic population. (Figure 2-3) The substrates in the mature biofilm are mainly transported into and out of the biofilm by diffusion (Figure 2-2). This means that the substrate concentrations typically decrease with depth in the biofilm and that the corresponding product concentrations increase with depth. (Figure 2-3) As a result non-favorable

conditions for the nitrifiers at the biofilm surface may become acceptable in the interior of the film if the easily biodegradable organics have been consumed already in the outer regions of the film. The thicker the biofilm the more likely there are operating conditions where the bacteria may coexist. Figure 2-3 shows the typical concentration profiles for DO, substrate BOD, $\text{NH}_4^+\text{-N}$, $\text{NO}_3^-\text{-N}$ and the thickness of various populations in biofilm. The relative concentrations of various populations, at any specific point in the treatment train, are a consequence of the nutrient supply, environmental conditions, and the degree of treatment received.

Figure 2-3 Concentration Profiles in Biofilm

(adopted from Watanabe et al., 1995)

2.10 Rate Control or Rate Limiting Conditions

In the process of removal of soluble organic matter, the organic matter serves as the electron donor and the oxygen serves as the electron acceptor. When nitrification is the objective, $\text{NH}_4^+\text{-N}$ serves as the electron donor and oxygen again serves as the electron acceptor. It is now widely recognized that in the most general case, the reactions within a biofilm may be controlled by the concentrations of both the electron donor and the electron acceptor. As with other biological processes, sufficient dissolved oxygen (DO) must be available in the wastewater within the RBC system to insure adequate treatment for BOD removal and nitrification. Availability of oxygen can be one of the major factors limiting the performance of an RBC. The rate of organic stabilization in RBC is generally limited by the oxygen penetration rather than by the substrate diffusion into the biological film especially in the initial stages. It is only in the multistage system that the substrate diffusion may become a rate-limiting factor in the last stages of the system. Wastewater DO levels of 1 to 2 mg/l are generally considered to be the minimum requirement to avoid DO limiting conditions.

With a strictly mechanically driven RBC, the only potential controls available to the operator are rotational speed adjustment and removable baffles. Increasing the disc rotating speed can increase the organic stabilization rate. This is because a higher rotating speed would cause greater oxygen transfer efficiency. Unfortunately, the rotating speed cannot be increased indefinitely without causing some major drawbacks. Firstly, power requirement increases with the disc rotating speed. Secondly, an excessively high rotating speed creates a high hydraulic shearing force, which may interfere with satisfactory development of biomass on the disc surface. Antonie, 1976; Antonie et al., 1974 have suggested that the optimal peripheral speed for treating domestic wastewater is about 18.2 m/min (60ft/min). Therefore, in order to increase oxygen penetration in RBC, some appropriate means other than unlimited increase in the rotating speed is necessary.

2.11 High Strength Wastewater and Associated Problems

For every RBC, treating either municipal or industrial wastewater, there is a limit to the rate at which the wastewater is applied beyond which first stage RBC units will not be able to maintain aerobic conditions. Table 2-3 shows the ideal ranges for the process parameters needed to be considered while designing RBC for the wastewater above 13°C. Much attention should be given to the first stage in design and operation, where organic loading should not exceed 30 g BOD/m² day or 12 g SBOD/m² day. Table 2-4 lists various studies carried out using high strength wastewater.

Table 2-3 Organic and Hydraulic Loadings Required for the Design of RBC Units (adopted from Tchobanoglous et al., 2003)

Parameter	Unit	Treatment		
		BOD Removal	Nitrification	BOD Removal and Nitrification
Hydraulic Loading	m ³ /m ² day	0.08-0.16	0.04-0.10	0.03-0.08
Organic Loading	g SBOD/m ² day	4-10	0.5-1.0	2.5-8.0
	g BOD/m ² day	8-20	1-2	5-16
Organic Loading (max - 1 st stage)	g SBOD/m ² day	12-15*	-	12-15*
	g BOD/m ² day	24-30	-	24-30
NH ₄ -N Loading	g N/m ² day	-	-	0.75-1.5
Hydraulic Residence Time	Hrs.	0.7-1.5	1.2-3.0	1.5-4.0

* WEF & ASCE, 1998

Table 2-4 Selected Cases of High Organic and/or Ammonia-Nitrogen Loading in RBC

References	Type of Wastewater	Loading					HRT
		Hydraulic X 10 ⁻³ m ³ /m ² .day	Organic		NH ₄ ⁺ -N		
			mgCOD/L	gCOD/m ² d	mg/L	gN/m ² d	
Alemzadeh et al. (2002)	Synthetic Phenolic	4-11	500-2000	22-52	-	-	Hrs 3.64-10
Apilnez et.al. (1998)	Synthetic Milk	157	730	115.26	0.18	0.284	2.93
Ayoub and Saikaly (2004)	Synthetic	93-95	800-1200	76-111.6	104-116	9.88-10.79	0.88-0.90
Banerjee (1997b)	Synthetic Phenolic	9-60	4-420 (Phenol)	1-28 (Phenol)	0.6-65	0.15-4.33	8.74-58.25
Brar and Gupta (2000)	Synthetic Trichloro ethylene (chlorinated solvent)	3.9-6.8 (30.9 shock loading)	1350 (COD) 600 (BOD)	5.27-9.18 (COD) 2.34-4.08 (BOD)	100	0.39-0.68	24-84
Gupta and Gupta (2001)	Synthetic High Strength Domestic Wastewater	6.66-10.66	500-1000	10-32	54.4-110.8	1-3.35	15-24
Israni et al. (2002)	Synthetic Phenolic	210.48-562.08	25-200 (Phenol)	42.10-84.19 (Phenol)	-	-	0.19-0.5
Kargi (2002)	Synthetic Saline	9.077-48	1500-14000	36.29-223.2	-	-	1.89-10
Kargi and Uygur (1997b)	Synthetic Saline	11.37-45.28	2400-11500	60-264	480-520	6-26.4	2-8
Radwan and Ramanujam (1997a)	Synthetic Phenolic (2-4 dichloro phenol)	24-65	671-1054	12.67-72.15	-	-	1.98-5.36
Radwan and Ramanujam (1997b)	Synthetic	30	1000-1557	29.99-46.7	20-185	0.66-5.5	4.2872
Saikaly and Ayoub (2003)	Synthetic Domestic	32-125	350-1200	11.03-111.6	45-116	1.42-10.79	0.68-2.47
Wilson (1997)	Synthetic High strength Pickling Waste	1.14-23	130-4800 (COD) 144-4449 (TOC)	1.7-43.8 (TOC)	-	-	4.97-100.25
Wilson and Wei (1997)	Synthetic	3-30	400-6500 (COD) 225-3700 (BOD)	1-21 (TOC)	-	-	3.13-31.33

Under heavy organic loads, DO profiles follow a pattern of rapid initial decline and slow recovery in successive stages, and the liquor from the first stage can be distinctly anaerobic. Low DO severely limits the oxygen transfer to the fixed film. In such an environment, the bacteria within the film interface are most likely anaerobic leaving only the bacteria on the outside surface of the interface aerobic. Once established this anaerobic condition there are a number of problems which can develop. Undesirable organisms will proliferate and retard oxidation of the BOD substrate. The two most common bacterial forms are anaerobic and sulfur oxidizing micro-organisms. The anaerobic bacteria normally co-exist with aerobic micro-organisms forming the underlayer of the biofilm (Figure 2-3). The aerobic layer is extremely thin and its depth is controlled in early stages by oxygen diffusion and in downstream stages by substrate diffusion.

The anaerobic underlayer is not inactive. The anaerobic bacteria can synthesize odor causing hydrogen sulfide from sulfate. When sulphide is present, either in the influent wastewater or by its production deep within the anaerobic biofilm, sulphide oxidising bacteria (SOB) such as *Beggiatoa* will thrive at the media interface. *Beggiatoa* utilize hydrogen sulfide and sulfur as energy sources in the presence of oxygen, according to the following equations (2-18) and (2-19) :

Although *Beggiatoa* exist under aerobic conditions, anaerobic conditions must be present for the formation of hydrogen sulfide and sulfur. Even with a correctly loaded system, *Beggiatoa* presents a potential problem if sulfide already exists in the system. *Beggiatoa* competes with heterotrophic organisms for oxygen and in extreme cases will take over the first stage of an overloaded RBC, shifting the load to the next stage and progressively taking over the system. Further, the growth is gelatinous, effectively blocking oxygen transfer and further inhibiting carbonaceous microorganism. When plenty of hydrogen sulfide is present, *Beggiatoa* store the sulfur in the cells, giving the organism the distinctive milky appearance (Frobisher, 1957). In addition to causing reduced BOD removal through the system, *Beggiatoa* growth has poor settling characteristics and will normally create a pinpoint white floc which easily crests the effluent weirs in the secondary clarifier. Thus, in the initial stages of a highly loaded RBC, large amounts of biomass exist, not contributing to the removal of organic materials and can stress the system. Overloaded RBCs performance could be improved by making minor process modifications (USEPA, 1984, 1993; WEF, 1992). Several methods that have been studied and proposed in the literature (Ito and Matsuo, 1980; Hittlebaugh and Miller, 1980; Kharajian et al., 1984; Brenner et al., 1984; Surampalli and Baumann, 1992, 1995, 1997; Surampalli et al., 1994; Kargi and

Uygur, 1997a; Saikaly and Ayoub, 2003; Ayoub and Saikaly, 2004) for improving the performance of rotating biological contactors (RBCs) in such situation include: (1) supplemental aeration; (2) step-feeding; (3) providing more disc area in first stage or enlarging first stage; and (4) recycling. Table 2-5 summarizes the selected studies carried out using RBCs with such process modifications.

Table 2-5 Selected Cases of RBC with Process Modification for Performance Improvement

References	Step Feeding	Recirculation	Enlarged First Stage	Supplemental Aeration
Ayoub and Saikaly (2004)	•	•		
Huang (1980)				•
Ito and Matsuo (1980)		•		
Janczukowicz and Klimiuk (1992)	•			
Kargi and Uygur (1997a)				•
Kevin et al. (1980)				•
Kharajian et al., 1984		•		
Klees and Silverstein (1992)		•		
Lagnese Jr. (1980)				•
Madden and Friedman (1980)				•
Neu (1994)		•		
Saikaly and Ayoub (2003)	•			
Surampalli and Baumann (1997), (1992)				•
Surampalli and Baumann, (1995)			•	•
Surampalli et al. (1994), (1984)				•
Tawfik et al. (2002)		•		

Methods other than supplemental aeration end up into increasing the number of stages. Increasing the number of stages will increase the total treatment potential of the system, reducing the stress. The

excess organic material left untreated by the first stage can be treated by the second and third stage, etc. Having several stages can dampen out organic slugs, but eventually results in more number of units and additional land requirement.

2.12 Diffused Aeration Systems and Oxygen Transfer Efficiency

For a given disc area and speed, the amount of oxygen transferred is fixed. An overstressed system will reduce the DO below the critical concentration and the efficiency of the system will be drastically reduced. It is not the relative disc size or oxygen transfer efficiency, but the absolute oxygen transfer capability of the system that determines whether the DO will be reduced below the critical concentration. A study (Albert, 1982) of a 36-shaft plant consisting of six rows of six shafts each, showed 96 % and 86 % removals of SBOD₅ and NH₃-N, respectively, when supplemental air was applied. The DO levels in all stages never dropped below 1.5 mg/l. The growth of *Beggiatoa*, a problematic filamentous bacteria, was eliminated with the exception of sparse growth on the media of the first stage. Prior to the addition of the supplemental air at that plant, the DO concentrations in the first four stages were measured at less than 1 mg/l and *Beggiatoa* growth was predominant. Without supplemental air, SBOD₅ removals of only 75 % were common and only about 60 % nitrification was achieved.

The addition of aeration would give the plant more flexibility in handling different types of wastes. Aeration may be the answer in situations where the waste characteristics are changed significantly by the addition of new industrial wastes after the plant has been constructed.

The two basic methods of aerating wastewater are (1) diffused-air aeration systems (2) mechanical aeration system. Diffused-air systems have been classified as either fine-bubble or coarse-bubble, with connotation that fine bubbles are more efficient in transferring oxygen. Fine-pore diffusers such as ceramic diffuser were the most efficient method of transferring oxygen, but unfortunately, they have a tendency to clog in wastewater within a year. So, surface aerators and coarse bubble diffusers were used even though the efficiency is lower. The newly developed 'Microporous Rubber Membrane Diffuser' (Figure 2-4) has solved this dilemma. This diffuser has the same excellent performance as fine-pore ceramic diffuser, and offers the following benefits over conventional diffusers.

- High oxygen transfer efficiencies
- Large interfacial surface area
- Even air distribution over the full range of air flows
- Low pressure drop
- High resistance to clogging and
- Back flow prevention.

Figure 2-4 Microporous Rubber Membrane Fine Bubble Diffuser

The capital cost of a fine-pore aeration system will probably be higher; however, the total annual cost for the system will be less than the annual cost of coarse-bubble aerators (USEPA, 1989). Higher oxygen transfer efficiencies of fine-pore diffusers compared to mechanical aerators and coarse bubble diffusers reduce the energy costs substantially. The fine-pore system could reduce energy consumption from 40 to 50 %, and overall life-cycle costs from 10 to 20 % compared to other diffused-air systems (USEPA, 1989).

The efficiency of oxygen transfer by diffusers depends on many factors including the type, size and shape of the diffuser; the airflow rate; the depth of submersion; tank geometry including the header and diffuser location and wastewater characteristics. Aeration diffusers are conventionally evaluated in clean water and the results adjusted to process operating conditions through widely used conversion factors. The chart in Appendix C - 1 shows the standard oxygen-transfer efficiency (SOTE) for the membrane diffuser used for this experiment. Typically, the SOTE increases with depth and the air flowrate. Standard Oxygen Transfer Rate (SOTR) can be calculated from Theoretical Oxygen Requirement (TOR) using SOTE by the following equation (2-20).

$$SOTR = \frac{TOR}{SOTE} \quad (2-20)$$

The conversion factors commonly used to convert the SOTR for clean water at test conditions to wastewater at treatment conditions, are α , β and θ .

The oxygen transfer correction factor α is used to estimate the $K_L a$ values for the actual system.

$$\alpha = \frac{K_L a (\text{wastewater})}{K_L a (\text{tap water})} \quad (2-21)$$

Where,

$K_L a$ Volumetric Oxygen mass-transfer coefficient, s^{-1}

Values of α vary with the type of aeration device, basin geometry, degree of mixing and the wastewater characteristics. Table 2-6 shows the values of α have been reported for fine-bubble diffuser systems (Hwang and Stenstrom, 1985) for various combinations of oxygen uptake rate (OUR) and airflow rate.

Table 2-6 Values of Oxygen Transfer Correction Factor α at Various Combinations of Airflow Rate and Oxygen Uptake Rate (Source: Hwang and Stenstrom, 1985)

Airflow Rate (m^3/hr)	Oxygen Uptake Rate (mg/L)		
	0	0.4	1.0
1.3	0.83	0.73	0.57
1.9	0.87	0.72	0.50
2.4	0.84	0.70	0.48
2.9	0.85	0.70	0.46
3.9	0.79	0.65	0.43

The oxygen uptake rate is related stoichiometrically to the organic utilization rate and growth rate, and can be defined as

$$r_o = -r_{su} - 1.42 r_g \quad (2-22)$$

Where,

r_o Oxygen Uptake Rate, mg O_2 /L.day

r_{su} Rate of Substrate Utilization, mg bsCOD/L.day

1.42 The COD of Cell Tissue, g bsCOD/ mg VSS

r_g Rate of Biomass Growth, mg VSS/L.day

The salinity-surface tension correction factor β , is used to correct the test-system oxygen transfer rate for differences in oxygen solubility due to constituents in the wastewater

$$\beta = \frac{C_w}{C_s}$$

(2-23)

Where,

C_w Operating oxygen concentration in wastewater, mg/L

C_s Oxygen saturation concentration for clean water at field-operating conditions, mg/L

Values of β vary from about 0.7 to 0.98. A β value of 0.95 is commonly used for wastewater (Tchobanoglous et al., 2003).

The temperature correction factor θ is used to correct the gas-liquid volumetric mass transfer coefficient, $K_L a$ value from test temperature to the treatment temperature, T . This is given by the Arrhenius equation (2-24).

$$K_L a_{(T)} = K_L a_{(20^\circ C)} \theta^{(T-20)}$$

(2-24)

Actual oxygen transfer rate (AOTR) can be calculated from the Standard Oxygen Transfer Rate using the correction factors by the following equation (2-25).

$$AOTR = SOTR (\alpha) \left(\frac{\beta C_s - C_w}{C_{s(20)}} \right) \theta^{(T-20)}$$

(2-25)

Where,

$AOTR$ Actual oxygen transfer rate under field-operating conditions, g O_2 / s

$SOTR$ Standardized oxygen transfer rate under test conditions at 20°C and zero dissolved oxygen, g O_2 / s

C_s Oxygen saturation concentration for clean water at field-operating conditions, mg/L

$C_{S(20)}$	Dissolved oxygen saturation concentration in clean water at 20 °C and 1 atm, mg/L
C_w	Operating oxygen concentration in wastewater, mg/L
T	Treatment temperature, °C
α, β, θ	Correction Factors

For diffused-air aeration systems C_s values must be corrected to account for the higher than atmospheric oxygen saturation concentrations achieved in the reactor due to release of air under a certain column of water pressure at the reactor bottom. The value of C_s is taken to be the average dissolved oxygen concentration attained at infinite time (ASCE, 1984)

2.13 Requisite Power for the Main Drive

Fujie et al. (1983) developed a relationship between the requisite power of RBC, P_w , and power number, N_p , by assuming the analogy between the disc assembly in and RBC and the impeller in a continuous stirred tank.

$$N_p = \frac{P_w g_c}{\rho N^3 D^5} \quad (2-26)$$

Where,

D	Diameter of the disc, m
ρ	Density of wastewater, kg/m ³
N	Speed of rotation, min ⁻¹
g_c	Gravitational conversion factor, kg m kg ⁻¹ min ⁻²

It should be noted that researcher considered the analogy between RBC discs with impeller in a continuous stirred tank which is usually completely submerged in the liquid, whereas RBC discs are usually 40 to 45% submerged.

Power Consumption at Low RPM:

Laminar flow condition in the liquid phase can be reasonably assumed at slow rotational speeds. Under laminar flow conditions, the power number was assumed to be inversely proportional to the disc Reynolds number based on the disc diameter and the distance between the discs

$$Np = \frac{\alpha'}{Re} \quad (2-27)$$

where α' is an empirical constant and the Reynolds number Re is given by the following equation, with the characteristic length d taken as the distance between contiguous discs.

$$Re = \frac{NDd\rho}{\mu} \quad (2-28)$$

Where, μ is viscosity of wastewater, kg/m s

Therefore, eliminating Np and Re by combining equations (2-26) (2-27) and (2-28) gives

$$P_w = \left(\frac{\alpha' \mu}{g_c d} \right) N^2 D^4 \quad (2-29)$$

The power requirement per unit surface area of discs is given by

$$\frac{P_w}{A} = \lambda_1 N^2 D^2, \quad \text{where } \lambda_1 = \frac{4\alpha' \mu}{\pi g_c d} \quad (2-30) \quad (2-31)$$

λ_1 (kW min² m⁻⁴) is an empirical constant, which depends on the distance between discs and the process conditions of the disc surface. The above expression is valid in the laminar regime only, up to a peripheral disc velocity of 0.5 m/s.

Power Consumption at High RPM:

At high rotational speeds, the liquid experiences turbulent flow rather than laminar flow. Under turbulent flow conditions, the disc power number can be considered to be independent of the Reynolds number. Under these conditions, the power requirements per unit surface area of discs is given by

$$\frac{P_w}{A} = \lambda_2 N^3 D^3, \quad \text{where } \lambda_2 = \frac{4N_p \rho}{\pi g_c}$$

(2-32) (2-33)

λ_2 is an empirically determined constant with dimension of $\text{kW min}^3 \text{m}^{-5}$.

Fujie et al. (1983) analyzed the data from various sources and determine the empirical values of constants λ_1 and λ_2 . (Refer Table 2-7)

Table 2-7 Empirical Values of Constants λ_1 and λ_2 (adopted from Fujie et al., 1983)

Disc Diameter (m)	$\lambda_1 \times 10^5$ (kW.min ² /m ⁴)	$\lambda_2 \times 10^5$ (kW.min ³ /m ⁵)		References
0.5	0.35	-	Smooth Disc	Sekine et al. (1979)
	1.10	-		
	1.75	0.45		
	1.10	0.20		
	1.30	0.26		
3.0	0.86	-		Antonie (1976)
1.75	0.39	0.52		USEPA (1971)

Sekine *et al.* (1979) obtained data under conditions where the microbial film was not attached to the disc surface. It should also be noticed here that the processing conditions of the disc surface affect the transient point from the laminar flow, to the turbulent flow region, and also affect the values of λ_1 and λ_2 .

Actual power (wired power) drawn by the main drive can be calculated by the expression

$$P_{w_{maindrive}} = \frac{P_w}{\eta_{maindrive gear} \eta_{maindrive motor}}$$

(2-34)

Where,

P_w Requisite power, kW

$P_{w_{maindrive}}$ Actual wired power drawn by main drive, kW

$\eta_{main\ drive\ gear}$ Overall efficiency of main drive reducing gear system

$\eta_{main\ drive\ motor}$ Efficiency of main drive motor

2.14 Requisite Power for Aeration

The following equation (Tchobanoglous et al., 2003) can be used to calculate the theoretical power requirement for a compressor operating adiabatically.

$$P_{adb} = \frac{wRT_1}{29.7n'} \left[\left(\frac{p_2}{p_1} \right)^{0.283} - 1 \right] \quad (2-35)$$

Actual wire power drawn can be calculated by the expression

$$P_{w_{blower}} = \frac{P_{adb}}{\eta_{blower} \eta_{blower\ motor}} \quad (2-36)$$

Where,

P_{adb}	Power requirement for adiabatic compression, kW
p_1	Absolute inlet pressure, atm
p_2	Absolute outlet pressure, atm
w	Mass flow rate of air, kg/s
R	Universal Gas Constant, 8.314 kJ/kmol °K
T_1	Absolute inlet temperature, °K
n'	$(k'-1) / k' = 0.283$ for air
k'	Constant = 1.395 for air

The overall efficiency for compressors is usually in the range of 0.7 to 0.9.

2.15 Definition of Power Economy

For activated sludge processes, 'Power Economy' is defined as the amount of oxygen dissolved into the water in the aeration tank per unit power consumption. This has been used as an indicator in evaluating the performance of a gas-liquid contacting process. Hosono et al. (1980) extended the concept of power economy to the trickling filter process. He defined the power economy as the amount of BOD removed per unit power consumption. The total amount of oxygen necessary to stabilize a waste is referred to as the oxygen demand.

The ultimate oxygen demand includes not only the amount of oxygen required to stabilize oxidizable carbonaceous materials, but also that which is required to microbially transform ammonia-nitrogen to nitrate-nitrogen. Hence, power economy could be defined in terms of ultimate oxygen demand (UOD) (Antonie, 1976) instead of 5 days BOD.

$$UOD = 1.5 BOD_5 + 4.5 NH_3 - N \quad (2-37)$$

Where NH_3-N is the ammonia content (mg/L) as nitrogen

2.16 Energy Consumption in RBC

Following the 70's oil crisis, energy use became a major concern in North America. Various sectors of US economy including wastewater treatment plants were singled out with regard to energy consumption. Consequently substantial research effort was directed towards the identification and development of energy-efficient treatment processes. The RBC process was identified as an energy efficient wastewater treatment process (Smith, 1973). Selected studies on energy consumption for RBC processes are listed in Table 2-8. Surprisingly the 70's publications on energy use in wastewater systems were based on theoretical calculations of energy consumption, not actual measurements (see Table 2-8)

Figure 2-5 shows a comparison of power consumption among various biodegradation processes used for the municipal wastewater treatments. The electric power consumption was measured in the existing plants with a daily flow rate of 300 m³/d. It shows that the RBC consumes much less energy than the other processes. Low energy consumption is one of the major reasons for the rapid growth in the use of the RBC process. Energy consumption in RBC treatment processes from some of the studies listed in Table 2-8 are summarized in Table 2-9.

Table 2-8 Selected Cases of Energy Consumption Studies about RBC and Comparison with Other Treatment Processes

Reference	Source of Data			Purpose of Study		
	Theoretical	Survey	Onsite/ Experimental	Comparison	Guideline Development	Benchmarking
Landon et al. (2003)			•	•	•	•
Benschine and Fok (2002)			•	•	•	•
Sauer and Kimber (2002)		•		•		
Kennedy (1997)	•			•		
Evans and Loughton (1994)		•		•		
Young and Koopman (1991)			•	•		
Owen (1982)	•			•		
Middlebrooks et al. (1981)	•			•		
Gernwood and Gloyna (1980)	•			•		
USEPA (1975b)	•			•		
Mills and Tchobanoglous (1974)	•			•		
Smith (1973)	•			•

Figure 2-5 Comparison of Energy Consumption among Various Biological Processes
 (Treatment capacity: 300 m³/day)
 (Source Watanabe and Iwasaki, 1997)

Table 2-9 Energy Consumption in RBC

Reference	Energy Consumption			Remarks
	Secondary Treatment Only		Total Plant	
	kWH/ kg BODr	kWH/m ³	kWH/m ³	
Kennedy (1997)	-	0.07 – 0.25	0.26 – 0.51	Generic
Landon et al. (2003)	-	0.28*	0.58	*Includes only main drives and aeration system
Benschine and Fok (2002)	1.1	0.17	0.28	
Sauer and Kimber (2002)	-	-	0.83@	@ Energy consumption for collection excluded
Evans and Laughton (1994)	-	0.20	-	
Young and Koopman (1991)	-	-	-	
Owen (1982)	0.93-1.70	-	-	kWH/kg BOD stabilized

The listed results are outcomes of either theoretical calculations based on electrical specifications of individual piece of equipment or data collected from existing treatment facilities through surveys / process parameter measurements. In the literature only conventional RBCs are studied for unit energy consumption. There are no published studies on unit energy consumption in modified RBCs.

3. EXPERIMENTAL SETUP

3.1 Introduction

Energy consumption per unit of oxygen demand by RBC processes depends on several operating and process parameters such as organic load, hydraulic load, rate of aeration, chemical characteristics of influent, degree of treatment required, environmental conditions, temperature etc. The objective of this study was to identify any opportunity of energy savings by balancing oxygen supply through fine bubble diffusers and rotating media. The optimization problem with many constraints and variables are simplified by keeping certain parameters constant and letting the remaining variables represent degrees of freedom.

In this chapter are described the experimental setup as well as the instrumentation required to determine the aforementioned parameters. Finally, it will be presented the variables and experimental matrix used in the present study.

3.2 Experimental Facility

The experimental facility shown in Figure 3-1 is located in the 'Laboratory of Water Treatment Technologies' at the Department of Chemical Engineering of Ryerson University. This facility is equipped with various commercial scale bioreactors such as a RBC (capacity 3000 L/day for domestic wastewater treatment), three-phase aerobic fluidized bed reactor and anaerobic fluidized bed reactor. The bioreactor used in this research was the rotating biological contactor. Table 3-1 lists various components of the experimental setup. Figure 3-2 shows the process flow diagram. Appendix A, summarizes the specifications of these equipment and instruments which were used for this research.

Sampling Points	
Tag	Description/ Location
1	Influent before Aeration Tank
A	Aeration Tank Exit
1	Stage 1 Exit
2	Stage 2 Exit
3	Stage 3 Exit
4	Stage 4 Exit
5	Stage 5 Exit
6	Stage 6 Exit

Rotating Biological Contactor –Flow Chart & Scheme for Experiment	
Prepared By : Rakesh Desai	Supervised By: Dr. M. Alvarez-Cuenca
MASc (2004-2006)	Ryerson University
Prepared Dt: Nov 11, 2004	Rev 04: Feb 08, 2006

Rev#	Date	Description
00	Nov11, 2004	Tanks Added
01	Jan 04, 2005	Modified for Offline Analysis
02	May 20, 2005	VFD added
03	Jun 02, 2005	Sampling Points; Rearrangement
04	Feb 08, 2006	

Figure 3-2 Rotating Biological Contactor –Flow Diagram & Experimental Setup

Table 3-1 List of equipment used in this study

	Equipment	Qty.	Purpose
Process Equipments	Vacuum Stripper	1	To strip off DO from tap water
	Feed Tanks	2	Synthetic wastewater preparation and supply
	Dosage Tank	1	Synthetic wastewater preparation
	Rotating Biological Contactor	1	Wastewater treatment, (Capacity 3000 L/d)
Accessories	Vacuum Pump	1	To generate vacuum in vacuum stripper
	Transfer Pump	1	To transfer water from vacuum stripper to RBC
	Main Drive Motor	1	Media rotation
	Reducing Gear (Main Drive)	1	To step down the speed
	Chain Drive (Main Drive)	1	To step down the speed
	Regenerative Blower	1	Source of air supply to the aeration tank
	Feed Pump	1	Influent pumping
	Fine Bubble Diffusers	6	Aeration
	Variable Frequency Drives	2	Speed control for main drive and aeration blower
	Metering Pump	1	Synthetic wastewater preparation
Measuring Instruments	Water Flow Totalizer	1	Synthetic wastewater preparation
	Water Flow Meter	1	Influent flow rate measurement
	Air Flow Meter	1	Airflow rate measurement
	Pressure Gauge	1	Blower discharge air pressure measurement
	Tachometer	1	Angular Velocity Measurement
	kWH Meter	2	Energy consumption measurement
Test Instruments	COD Reactor	1	COD sample digestion
	Photometer	1	COD test
	pH meter with electrode	1	pH measurement
	LDO meter with electrode	1	DO & temperature measurement
	ISE meter	1	$\text{NH}_4^+\text{-N}$ & $\text{NO}_3^-\text{-N}$ measurement
	Ammonia Electrode	1	$\text{NH}_4^+\text{-N}$ measurement
	Nitrate Electrode	1	$\text{NO}_3^-\text{-N}$ measurement

3.3 Details of Process Equipment

The vacuum stripper, a feed tank, a modified RBC with fine bubble aeration system, and various instruments to measure energy consumption and the synthetic wastewater constituents were the major components used in for this study.

3.3.1 Process Equipments

A. Vacuum Stripper

Figure 3-3 Vacuum Stripper Used for Deaeration of Tap Water

The experiment was carried out in two phases. The RBC was operated with tap water and without any biofilm during phase-I study. In phase-II of the experiment, the modified RBC was operated with fully developed biofilm on media. The vacuum stripper shown in Figure 3-3, was used to remove dissolved oxygen from tap water before feeding it to the rotating biological contactor. The stripping of oxygen and subsequent oxygen transfer in the reactors permit the evaluation of oxygen mass transfer parameters. Appendix A (A) explains the characteristics of this vacuum stripper. Vacuum stripper is a packed column made of polyvinyl chloride (PVC) and consists of the following four zones:

- Demister,
- Two packing stages (includes water distribution / redistribution zone), and
- Reservoir for deoxygenated water storage and supply

Figure 3-4 Tri-Packs® - Packings Used in Vacuum Stripper¹

Each stage consists of a section of Tri-Packs®, hollow spherical column packing (see Figure 3-4). Nesting is virtually impossible with this packing and there is no need for allowances for settling. Spherical shaped packing roll themselves into the "packed position" without forming void areas common to irregular shaped packing or those with excessive pins and appurtenances. The unique network of ribs, struts and drip rods provide a distinct advantage of excellent wetting qualities and maintain proper liquid distribution through the packed bed. Excessive surface area can impede proper gas and liquid contact and always results in higher pressure drop and increased operating costs. An optimum surface area to open area ratio of Tri-Packs® yields excellent mass transfer efficiency and reduced operating cost.

¹ Tripacks® – Jaeger Products Inc., 1611 Peach leaf St., Huston, TX 77039 Tel (281) 449-9500; 1-800-678-0345, www.jaeger.com

The solubility of any gas in a liquid is inversely proportional to the temperature and directly proportional to the pressure. Hence, reducing the absolute pressure at constant temperature can result in decreased solubility of dissolved oxygen in water. An oil ring vacuum pump (Figure 3-5) was used for the purpose of reducing absolute pressure in the vacuum stripper. The vacuum pump is connected to the top of the vacuum stripper, and was run continuously during the study in phase-I.

Figure 3-5 Vacuum Pump Connected to the Vacuum Stripper

Tap water was evenly distributed through sixteen water-sprinklers in the form of fine droplets over the upper packed bed. Sprinkled water was trickled down through the packed bed in the form of thin film on the packing, which provided very large surface area for efficient mass transfer. In order to avoid channeling, the packed bed is divided into two stages instead of one continuous bed. Dripping water from the upper bed is redistributed over the lower packed bed. After passing through the lower packed bed, deaerated water was collected into the reservoir at the bottom of the vacuum stripper. The vacuum pump is capable of developing 27-28 inHg of vacuum in the vacuum stripper, and in turn achieving less than 1.5 mg/L of dissolved oxygen concentration in tap water. In order to prevent deterioration and to maintain the reliable performance of the vacuum pump, suspended fine water droplets from air was collected by sucking air through a demister. The Demister is a 20 cm thick bed of plastic wool at the top of the vacuum stripper.

Figure 3-6 Transfer Pump to Feed Deaerated Water to the RBC

The transfer pump (Figure 3-6) is a progressive cavity pump, which was used to feed the deaerated water to the modified RBC unit. This pump has a very high capacity compared to the requirement of deaerated water during phase-I. This required recirculation of excess water back to the vacuum stripper, which was achieved by adjusting the delivery valve, recirculation valve and RBC's feed valve.

B. The Feed Tank

The feed tank shown in Figure 3-7 (A), was used to prepare, store and supply synthetic wastewater for the study in phase-II. The feed tank is a high-density polyethylene vertical cylindrical tank with a capacity of 5000 Liters. This tank was capable of holding three days supply of wastewater for this study. As the same feed tank was supplying wastewater for other experiments simultaneously, it was required to prepare fresh synthetic wastewater at least once a day or even more frequently.

Figure 3-7 Feed Tanks Used to Prepare and Store Synthetic Wastewater

This 5000 Liter feed tank was replaced by a 160 Liter square section tank (Figure 3-7 -B) after facing operational problems during the seeding process, due to undesirable biological activity in the former tank.

C. The Rotating Biological Contactor

The modified RBC shown in Figure 3-8 was used to evaluate oxygen mass transfer, removal performance and effectiveness of oxygen supply through fine bubble aeration and rotating media in order to minimize unit energy consumption. The treatment unit used in this study was a commercial full-scale unit, which can treat 3000 L/day of domestic wastewater. The unit consists of an aeration chamber, a multistage rotating disc treatment system and a clarification tank. The aeration chamber, RBC cradle and clarifier are molded using fiberglass reinforced plastic (FRP) in a single piece. The features of modified RBC are summarized in Appendix A (C).

Figure 3-8 Modified Rotating Biological Contactor Structure and Components

Aeration Chamber

Deaerated tap water in phase-I study and synthetic wastewater in phase-II study were fed to the aeration chamber (see Figure 3-8) using a progressive cavity feed pumps. The aeration chamber was used to aerate the influent wastewater before flowing through the stages of the RBC. This was used to

decrease the organic load on first stage and to improve DO concentration throughout the unit. High efficiency fine bubble membrane diffusers (Appendix A) were used for aeration. The characteristics of the air diffusers are described in Appendix A (I). Air was supplied using a regenerative blower.

Figure 3-9 Diagram of a Fine bubble diffuser Used in the Aeration Chamber
(adopted from www.diffuserexpress.com)²

Disc Assembly

The disc assembly consists of six stages in series (Figure 3-8) to achieve simultaneous carbonaceous BOD and ammonia oxydation. The RBC cradle is constructed of FRP and divided into six stages using semi-circular FRP partitions along the length. The support media discs are fabricated from light-weight polyethylene mesh (Figure 3-10), and the discs are mounted on a horizontal rotating shaft. The discs assembly shown in Figure 3-10 has a surface area density of approximately 37 ft²/ft³ and is 40 % submerged in the wastewater. After passing through the aeration chamber, the wastewater was flowing through stages of the rotating disc treatment system in series. The wastewater was flowing from one stage to the next via notches in the partitions. The longer residence time of the influent, zigzag arrangement of notches and turbulence produced by the rotating discs created the homogeneous condition in each stage.

Figure 3-10 Disc Media

² Environmental Dynamics Inc. 5601 Paris Rd., Columbia, MO 65202, Tel: 573-474-9456, www.diffuserexpress.com

Clarifier

The sludge is settled by gravity in the final clarifier and the clear effluent treated overflows to the sewer for disposal. The clarifier is an integral part of the RBC unit.

Modifications

In order to make the existing equipment suitable for the study, the following modifications were carried out.

1. Removal of the partition separating the Aeration Chamber from the Primary Clarification Tank

The RBC used for this study had a partition (Figure 3-11 - before), separating the aeration chamber from the primary clarification tank. The volume occupied by the primary clarification tank was 0.91 m^3 leaving only 0.35 m^3 for aeration chamber. This partition was removed to utilize the combined tank as an aeration chamber (see Figure 3-11).

Before

After

Figure 3-11 Aeration Chamber Before and After Removal of the Partition

Before

After

Figure 3-12 Aeration Diffuser Used in RBC's Aeration Chamber (Before and After)

3. EXPERIMENTAL SETUP

Rev#	Date	Description
00	Dec 01, 2005	

Diffuser Layout in the Rotating Biological Contactor's Aeration Chamber		
Prepared By : Rakesh Desai	Supervised By: Dr. M. Alvarez-Cuenca	Prepared Dt: Dec 01, 2005
MA Sc (2004-2006)	Ryerson University	Rev 00: Dec 01, 2005

Figure 3-13 Diffuser Layout in the Rotating Biological Contactor's Aeration Chamber

2. Replacement of the Coarse Air Diffuser by the Fine Bubble Air Diffusers

The aeration chamber had a coarse bubble air diffuser (Figure 3-12-before). This coarse bubble diffuser was replaced by six high efficiency 9" Φ fine bubble diffusers (Figure 3-12-after) to achieve efficient aeration. The calculation of the number of diffusers required for the aeration is summarized in Appendix B. Figure 3-13 shows the diffuser layout in the aeration tank.

3. Replacement of the Central Compressed Air Supply by the Aeration Blower

The former experimental setup utilized central compressed air supply for aeration. Furthermore, this research seeks to collect data on energy consumed during the treatment of wastewater. Hence, the regenerative air blower (Figure 3-14) dedicated to the RBC replaced the compressed air supply. Appendix B, shows blower sizing calculation. The calculation (Appendix B) indicates that the selected blower has the capability of supplying at least 15 SCFM air at 60 inWG static head. The blower 'SCL V3' manufactured by FPZ was selected, as it was capable of supplying 20 SCFM at 60 inWG static head.

Figure 3-14 Blower³ to Supply Air to the Aeration Chamber

4. Replacement of the Mono-Block System by 3-phase Motor and the Variable Frequency Drive

The RBC had initially a mechanical variable gear system attached to the main drive (Figure 3-16), to control the rotational speed of the disc assembly. This mono-block drive system was evaluated for the variation of power consumption with speeds. Table 3-2 shows the amperage drawn by the main drive at various angular velocities of the discs. Statistical analysis of the data (Table 3-3) did not show any

³ FPZ Inc., 2091 N. Port Washington Rd, Grafton, WI 53024-9747 Tel: (262) 375-7360, Web: www.FPZ.com

considerable variation in current drawn with changes in speed. The correlation between angular velocity and the amperage was very poor (Figure 3-15). This phenomenon can be associated to the friction losses in the gear system. On the basis of these results, the mono-block main drive system was replaced by a combination of 3-phase AC motor and variable frequency drive (Figure 3-16).

Table 3-2 Current Drawn by Main Drive at Various Angular Velocity of Discs

Main Drive Motor: 1/2 HP; 60 Hz; 115 V; 1 Phase; 1725 RPM
 Reduction Gear: 'Varitrac' - Speed Range 345-2060 RPM
 Ammeter: 'SPERRY DSA-400, Digisnap' Digital Snap-around Volt-Ohm-Ammeter (20,200,400A)

Speed Setting	RPM	Current Drawn (Amp)		Average	Variance
		Increasing Speed	Decreasing Speed		
1	1.18	8.045	7.900	7.973	0.0105
2	1.57	7.990	7.905	7.948	0.0036
3	1.97	7.975	7.875	7.925	0.0050
4	2.47	7.935	7.905	7.920	0.0005
5	2.99	7.980	7.885	7.933	0.0045
6	3.51	7.910	7.885	7.898	0.0003
7	3.98	7.950	7.930	7.940	0.0002
8	4.53	7.935	7.885	7.910	0.0012
9	4.98	7.925	7.885	7.905	0.0008
10	5.51	7.885	7.845	7.865	0.0008
11	6.00	7.825	7.850	7.838	0.0003
12	6.53	7.895	7.885	7.890	0.0001
13	6.99	7.925	7.895	7.910	0.0005
14	7.38	7.905	7.96	7.933	0.0015
Sum	59.59	111.08	110.49		
Avg	4.26	7.93	7.89		
Var	4.230	0.0028	0.0008		
SDev	2.057	0.0532	0.0291		

Table 3-3 Analysis of Variance for Current Drawn by Main Drive at Various Speeds

ANOVA: Single Factor with replication						
Source of Variation	Sum of Squares	Degree of Freedom	Mean Square Value	F	$P_{\alpha_{max}}$	$F_{critical}$
Between Speeds	0.030536	13	0.002349	1.104437	0.426267	2.507264
Within Groups	0.029775	14	0.002127			
Total	0.060311	27				

- Null Hypothesis H_0 : Current drawn at various speeds are the same
- $F_{13;14;\alpha_{max}} \ll F_{13;14;5\%}$ (F critical) hence $\alpha_{max} \gg 5\%$
- Or alternatively $P_{\alpha_{max}} = 42.6267\%$ ($\ll 85\%$)
- Thus, H_0 (Null Hypothesis) must be accepted and it can be concluded that the current drawn at various speeds are the same

Figure 3-15 Correlation between Main Drive (Previous) Speed and Current Drawn **Before** **After** **Variable Frequency Drive****Figure 3-16 Main Drive System Before and After Modification**

3.4 Experimental Matrix

This study was carried out in two phases. Phase-I without biofilm and phase-II with biofilm developed. To study the independent effect of the angular velocity of the disc assembly and the aeration rate on steady-state oxygen mass transfer, the RBC was operated with tap water and without any biofilm developed on the media during phase-I. In phase-II of the experiment, performance and energy consumption in the modified RBC were studied after biofilm was developed on the discs and analysis of the effluent was at steady state. In this section, constant, independent and dependent parameters are listed and the experimental matrices for both phases are discussed.

Constant Parameters

Parameters other than those related to the rate of oxygen supply; i.e. hydraulic load, organic load, influent chemical characteristics, temperature etc. were relatively kept constant throughout this study. Table 3-4 lists some of the significant constant parameters.

Table 3-4 Constant Parameters

Influent Flow Rate	Influent COD	Influent $\text{NH}_4^+\text{-N}$
1500 L/day	5000 mg/L	200 mg/L

Independent Variables

The rate of oxygen supply to the RBC trough and aeration chamber are the parameters that were directly linked to the oxygen demand removal and hence to the unit energy consumption. These parameters were varied by controlling the main drive speed and the aeration blower speed respectively, using variable frequency drives. Table 3-5 summarizes the operating ranges for these variables. The configurations for the experiments without and with biofilm are summarized in Table 3-6 and Table 3-7 respectively. Each combination of independent parameters was run only once without duplication.

Table 3-5 Independent Variables

Rotor RPM	Blower Speed/ Flow Rate
2.5-5 rpm	0-15 CFM

Table 3-6 Experimental Matrix (Without Biofilm)

Air Flow Rate CFM	Angular Velocity of Discs (RPM)			
	2.5	3.0	4.0	5.0
0	Run (5)	Run (11)	Run (7)	Run (14)
5	Run (9)	Run (15)	Run (13)	Run (1)
10	Run (6)	Run (10)	Run (2)	Run (12)
15	Run (16)	Run (4)	Run (8)	Run (3)

Table 3-7 Experimental Matrix (With Biofilm)

Air Flow Rate CFM	Angular Velocity of Discs (RPM)			
	2.5	3.0	4.0	5.0
0	Run (16)	Run (15)	Run (14)	Run (13)
5		Run (9)	Run (10)	
8			Run (11)	
10	Run (8)	Run (7)	Run (6)	Run (5)
12			Run (12)	
15	Run (1)	Run (2)	Run (3)	Run (4)

Dependent Parameters

Table 3-8 lists dependent parameters which were critical to the study of the unit energy consumption and performance of the RBC.

Table 3-8 Dependent Parameters

Wastewater Characteristics (in-process)				Power Consumption (kWH)	
DO mg/L	COD mg/L	NH ₄ ⁺ -N mg/L	Nitrate-N mg/L	Main Drive	Blower

3.5 Study of Energy Consumption and Oxygenation Capacity of RBC without Biofilm (Phase-I)

3.5.1 Energy Consumption by the Main Drive and Aeration Blower in Absence of Biomass

A main drive system includes various components such as a VFD, an electric motor, a worm gear and a chain drive. The theoretical determination of the energy consumptions by the rotating media and the aeration blower were not possible as efficiencies of these components are either difficult to determine and/or vary with speed. Hence, before starting the experimental runs using synthetic wastewater, energy consumption by the main drive without biofilm was studied. The RBC unit was completely filled up with tap water. The VFD was set at different frequencies along the full range. The RBC was run for at least one day for each speed setting. Angular velocities, various electrical parameters and energy consumption were logged using the form shown in Appendix F - 1. Appendix D - 1 shows the correlation developed during this study, between % frequency set at VFD with the angular velocity of the disc assembly, amperage and power. Similarly, power consumption by the aeration blower was also studied with the aeration chamber filled with tap water. Appendix F - 2 was used to record the rate of aeration, static air pressure, various electrical parameters and energy consumption. Appendix D - 2 shows the correlations developed from the experimental results, between % frequency set at VFD with rate of aeration, static air pressure, amperage and power drawn by the blower.

3.5.2 Steady State DO Profile Study

In phase-I, was studied the oxygen transfer capability of the RBC in the absence of biodegradation process. In this phase, were developed steady-state dissolved oxygen profiles. This was carried out in the modified RBC at various combinations of angular velocity and aeration rates.

The vacuum pump connected to the vacuum stripper was switched 'ON'. After the pressure in the vacuum column reached 0.06 atm to 0.1 atm, the tap water valve feeding the vacuum stripper was opened. The tap water fed to the stripper was sprinkled over the top of the upper stage through sixteen water sprinklers. The temperature of the feed water was maintained at 15 ± 2 °C. The water passing through the two stages of packing was deaerated and collected into the reservoir at the bottom of the vacuum stripper. After the water level in the reservoir reached 50% of its capacity, the transfer pump was switched 'ON'. The transfer pump has a capacity larger than the required feed rate. The recirculation valve was adjusted to make the water flow rate to the RBC a little higher than the required 1 L/min. This was necessary to avoid excessive pressure build-up in the feed pipe. The recirculation valve allowed the remaining deaerated water back to the vacuum stripper. The feed valve at the RBC was adjusted to achieve a deaerated water flow rate of 1 L/min. The tap water feed valve was adjusted so that the water level in the reservoir will be maintained between two level switches.

The RBC cradle and the aeration chamber were emptied and cleaned before the phase-I study. The RBC unit was completely filled up with deaerated water. After filling-up the RBC unit completely, the main drive was turned 'ON' and the Variable Frequency Drive (VFD) was adjusted to achieve desirable speed. The aeration blower was turned 'ON' when required and the VFD was adjusted to achieve the desirable airflow rate. The experimental runs were carried out in the sequence mentioned in Table 3-6.

3.6 Preparation of Synthetic Wastewater

Sewage is a complex mixture of many inorganic and organic substances including microorganisms and varies its composition with time. To make reproducible the present experimental work it was decided to produce synthetic wastewater. Thus, high strength wastewater was made using tap water and predetermined dosages of sugar and commercial grade inorganic chemicals. A combination of hot and cold tap water was instrumental to keep the temperature of the synthetic wastewater relatively constant and at room temperature. Chemicals from the same production lots were used to eliminate variations in chemical composition and amount of impurities present. Table 3-9 shows the chemical composition of the synthetic wastewater.

Table 3-9 Composition of the Synthetic Wastewater Used

Compound	Concentration mg/L	mg/L	
Sugar	3233	5000 COD	1
(NH ₄) ₂ SO ₄	943	200 N	
KH ₂ PO ₄	56	12.75 P	2, 3
K ₂ HPO ₄	144	25.50 P	
MgSO ₄ - 7H ₂ O	50		4
FeSO ₄ - 7H ₂ O	4		4
CaCl ₂ - 2H ₂ O	30		4

1. For sugar solution- BOD₅ ≈ 60-70% COD (Soft drink Plant Wastewater Processing Using Reverse Osmosis (www.gewater.com/library/tp/722_Soft_Drink.jsp))
2. Calculated based on C:N:P Ratio 100:5:1 (Gray, 2004; Pufal et al., 2003; Jong-Hyun Lee, 1999; Mimma Pemetti, 2003)
3. Ratio of P from KH₂PO₄ : P from K₂HPO₄ was maintained at 1:2 ratio
4. BOD₅ test method (APHA 1995)

Sugar and ammonium sulphate were used as a source of BOD₅ and NH₄⁺-N respectively. Wastewater can only be treated biologically if sufficient nitrogen and phosphorus are present. The optimum C:N:P (BOD₅:N:P) weight ratio is 100:5:1 (Gray, 2004). In the present study, C:N:P (BOD₅:N:P) weight ratio was maintained approximately at 81:5:1. This ratio was selected to closely match the ideal ratio and maintain BOD₅ and NH₃-N around 3000 mg/L and 200 mg/L respectively. A combination of potassium

dihydrogen phosphate (KH_2PO_4) and potassium hydrogen phosphate (K_2HPO_4) was used both to buffer the mixed liquor and to provide a phosphorus source. P from KH_2PO_4 to P from K_2HPO_4 ratio was maintained at 1:2. This ratio was selected based on the standard method for the BOD_5 test (APHA 1995). The influent pH was maintained between 6 and 6.3 using sodium hydroxide (NaOH) when required. Besides these components, iron in the form of $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}$ and magnesium in the form of $\text{MgSO}_4 \cdot 7\text{H}_2\text{O}$ were added.

The list of chemicals in Table 3-9, does not include trace elements. Trace elements are required by certain microorganisms in very small amounts and are naturally present as 'contaminants' of the water. In addition, they are also available as 'impurities' in commercial grade chemicals. Hence it was not necessary to add them to the synthetic wastewater. Table 3-10 shows the characteristics of the resultant synthetic wastewater.

Table 3-10 Characteristics of the Synthetic Wastewater

Parameter	Concentration	
COD	5000	mg/L
$\text{NH}_4^+\text{-N}$	200	mg/L
pH	6-6.3	-
Temperature	22-30	°C

Figure 3-17 shows the process diagram for the preparation of the synthetic wastewater. Because of undesirable biological activity synthetic wastewater was required to be freshly prepared and immediately fed to the bioractors of the laboratory.

'Setup A':

Figure 3-17 ('Setup A') shows the original setup for synthetic wastewater preparation in which the 5000 Liter feed tank was used for the preparation of the synthetic wastewater. A 160 liter, high-density polyethylene dosage tank (Figure 3-18) was used to prepare a concentrated solution of sugar and commercial chemicals. Concentrated synthetic wastewater and tap water were fed to the feed tank simultaneously. The flow rate of the tap water and concentrated synthetic wastewater was adjusted to complete the preparation in approximately 2 hours. While feeding to the RBC, recirculation of excess influent kept the synthetic wastewater properly mixed.

3. EXPERIMENTAL SETUP

Sr	Symbol	Description
1.		Metering Pump
2.		Water Meter
3.		Ball Valve
4.		Diaphragm Valve
5.		Measuring / Analysis

Setup A

Setup B

Rev#	Date	Description
00	Sep14, 2005	
01	Feb 04, 2006	Metering Pump Added
02	Mar 12, 2006	Ball an Pin valves removed
03	Jun 28, 2006	Setup B added

Scheme for Synthetic Wastewater Preparation

Prepared By : Rakesh Desai	Supervised By: Dr. M. Alvarez-Cuenca	Prepared Dt: Sep 14, 2005
MASc (2004-2006)	Ryerson University	Rev: 03 Jun 28, 2006

Figure 3-17 Schematic for the Preparation of Synthetic Wastewater

Figure 3-18 Dosage Tank and Metering Pump Used in the Preparation of Synthetic Wastewater

Setup B:

The initial operation of the experimental assembly revealed difficulties maintaining the pH and preventing undesirable bioactivity in the feed tank during the seeding process. Therefore, the 5000 liter feed tank was replaced by a 160 liter square section tank. (Setup B - Figure 3-17). This square section tank (Figure 3-7 - B) was used for real time mixing of the concentrated synthetic wastewater and the tap water before feeding the bioreactors. Chemicals and sugar for 8 hours of supply (≈ 1000 L of synthetic wastewater) were mixed in 96 liter tap water in the dosage tank (Figure 3-18). The concentrated solution was continuously fed to the square section feed tank using a metering pump (Figure 3-18) at the rate of 0.2 L/min. Simultaneously, tap water was fed to this square section tank at the rate 1.8 L/min. The RBC and the Fluidized Bed Reactor (FBR) were continuously fed each at a 1 L/min flow rate.

3.7 RBC Start-up, Seeding and Operation

The experimental operation demands the creation of biofilm on the discs. This was achieved by seeding the synthetic wastewater with water and sludge brought from the Stamford Niagara Pollution Control RBC Plant in Niagara Falls, Ontario. A total of 45 L of seed microorganisms were collected from the primary, secondary and 1 stage effluents of the RBC plant of the aforementioned municipal wastewater treatment plant. The RBC used in present study was filled up with the synthetic wastewater and the 45 L mixture of seed wastewater was then distributed among the aeration chamber and the six RBC stages.

The main drive and the aeration blower were then switched 'ON'. The angular velocity of the discs was set at 0.8 RPM and the rate of aeration was set at 3 SCFM. The reactor was operated in a batch mode,

to provide the initial colonization and accumulation of microorganisms. The RBC was operated without any flow of wastewater for a day. On the second day, the progressive cavity feed pump was switched 'ON' and the wastewater recirculation was started to speed up the growth process. The recirculation rate was set at 1 L/min. The angular velocity of the discs and the airflow rate in the aeration chamber were increased to 1 RPM and 5 SCFM respectively. The RBC was operated in the batch mode with a recirculation rate of 1-4 L/min till day 7. During the first week, daily 100-200 Liters of fresh synthetic wastewater was fed to the RBC to makeup the evaporation losses and to maintain sufficient organic load for proper growth. Microorganisms present in the seed water started to adapt to the new environment which resulted into the thin biofilm formation on the media and suspended growth in the aeration chamber. No pH adjustment was performed. Within three days of the operation, a transparent biofilm was clearly visible in the first three stages. On the 8th day, the biofilm grew relatively thick in the two initial stages and very thin in the two final stages of the RBC. On the 8th day, the RBC was transferred from a batch mode to a continuous mode. The feed rate was set at 0.1 L/min. During the second and third week of operation, the wastewater feed rate and angular velocity of the discs were gradually increased to 0.68 L/min and 3 RPM respectively. This resulted into fully developed suspended growth in the aeration chamber and biofilm on the discs. Whole seeding and startup process took about 45 days. The feed rate, angular velocity and rate of aeration were set to 1 L/min, 2.5 RPM and 15 SCFM respectively. The progress and problems that arose during the seeding process are further discussed in Section 4.2. Once steady state was reached, a total of sixteen experimental runs were conducted under varying operating modes (Table 3-7).

3.8 Experimental Runs (Phase-II)

Throughout the study in phase-II, the synthetic wastewater was fed continuously into the aeration chamber of the RBC using a progressive cavity feed pump at a rate of 1500 L/day (1 L/min). Air demand in the aeration chamber was catered through fine bubble air diffusers using a regenerative air blower dedicated to this experimental setup. The regenerative blower speed was controlled by setting a variable frequency drive to achieve different rates of aeration required for various experimental runs (Table 3-7).

After leaving the aeration chamber, partly biodegraded wastewater entered into the first stage and then moved through the six stages in series. The variable frequency drive was set to control the main drive speed in order to achieve various angular velocities of the discs required for the experimental runs (Table 3-7). From the sixth stage of the RBC, mixed liquor entered the secondary clarifier. Upon leaving the clarifier, the treated clear effluent was disposed into the sewer. The settled sludge from the secondary clarifier was removed once every week. This was done by mechanically mixing wastewater and sludge using air and adding fresh water to replace mixed liquor,

Influent feed rate and composition were kept constant throughout the study as shown in Table 3-4, Table 3-9 and Table 3-10. Table 3-11 summarizes the operating parameters for the experiment. Steady state condition for each run was ensured by measuring DO and pH at each sampling point. Steady-state was presumed to have been established when three successive set of results were within 5%.

Table 3-11 Experimental Operating Parameters

		Loadings				
Flow Rate, m ³ /day		1.5				
Hydraulic Loadings, m ³ /m ² day		0.017				
BOD ₅ , mg/L (kg/day)		3250 (4.875)*				
Ammonia – N, mg/L (kg/day)		200 (0.30)				
Organic Loadings (BOD ₅), gBOD/m ² day		55.970				
Ammonia - N Loadings, g/m ² day		3.444				
Hydraulic Residence Time						
		Volume (m ³)	Residence Time (hrs)			
Stage 1		0.180	2.88			
Stage 2		0.100	1.60			
Stage 3		0.100	1.60			
Stage 4		0.100	1.60			
Stage 5		0.120	1.92			
Stage 6		0.115	1.84			
Total (Stages)		0.715	11.44			
Aeration Tank		1.260	20.16			
Sedimentation Tank		0.610	9.76			
Total Unit		2.585	41.36			
Peripheral (Tip) Speed at Different RPM						
RPM	1	2	3	4	5	6
Tip Speed (m/s)	0.0461	0.0922	0.1382	0.1843	0.2304	0.2765

* Assumed 65 % of COD

3.9 Measurement of Process Parameters

Various measuring instruments listed in Table 3-1 and shown in Figure 3-19, were used to measure various process parameters. Locations for these measuring instruments are indicated on process flow diagrams (Figure 3-2) and the synthetic wastewater preparation diagram (Figure 3-17). Specifications of these measuring instruments are summarized in Appendix A (L-Q).

Water Flow Totalizer

Airflow Meter

Water Flow Meter

Air Pressure Gauge

kWH Meter

Tachometer

Figure 3-19 Instruments Used to Measure Various Process Parameters

Table 3-12 lists the measuring instruments used and also indicates the frequency of measurement. Observations for these parameters were recorded daily and/or once after the steady state was reached.

Table 3-12 Process Parameter Measurement Protocol

Process Parameter	Measuring Instrument	Frequency
Vacuum	Vacuum Gauge	At least hourly while vacuum pump is running
Total Volume (Water)	Water Flow Totalizer	During synthetic wastewater preparation
Temperature	Built-in temperature sensor in DO meter	At least once a day
Wastewater Feed Rate	Rotameter	At least once a day
Angular Velocity	Tachometer	At least once a day
Air Flow Rate	Rotameter	At least once a day
Air Pressure	Burden Pressure Gauge	Daily
Energy Consumption (Main Drive)	kWH meter	After Steady State reached
Energy Consumption (Blower)	kWH meter	After Steady State reached

3.10 Analytical Equipment

The equipment/instrumentation used in the analytical test program are listed in Table 3-1. Specifications of this test equipment are summarized in Appendix A (R-X). A subsequent discussion gives the details and the features of the test equipment used.

A. Dissolved Oxygen Measurement:

Dissolved oxygen concentrations were measured using the Hach LDO™ (Luminescent Dissolved Oxygen) HQ10 Portable Dissolved Oxygen Meter (Figure 3-20).

Figure 3-20 Portable LDO Meter with Probe for DO measurement
(adopted from www.hach.com) ⁴

⁴ Hach Company, 100 Dayton Avenue, Ames, Iowa 50010 Telephone: (515) 232-2533

The LDO™ measurement principle is based on the physical appearance of luminescence. This is defined as the property of certain materials (luminophore) to emit light as a result of excitation by light. A thin film of platinum based luminescent material called luminophore is entrapped in an oxygen permeable polymer. Blue light from an LED is transmitted to the sensor surface. The blue light excites the luminescent material. Upon relaxation (dissipation of excitation energy), light of lower energy i.e. longer wavelength, the red light is emitted (luminescence). The time from when the blue light was sent and the red light emitted is measured. The more oxygen present, the shorter the time it takes for the red light to be emitted. This time is measured and correlated to the oxygen concentration. Between the flashes of blue light a red LED is flashed on the sensor. This internal red LED provides a reference measurement before every reading to ensure the sensor's accuracy is maintained. The Hach LDO® probe offers the following key advantages over other meters that are based on galvanic or polarographic sensor technology:

- No membranes and electrolyte to replace, and no cathodes/anodes to polish
- No interference from typical wastewater chemicals
- Pre-calibrated sensor, no user-calibration required for the life of the sensor
- No polarization time
- Fast response, with minimum to no stirring required
- Durable sensor retains accuracy even when scratched or partially fouled and can easily be cleaned

Interferences

There are no known interferences at normal wastewater concentrations. Chlorine dioxide is the only known interference, and may give high readings. The use of synthetic wastewater ensures the absence of chlorine dioxide and the reliability of the results.

Maintenance

The LDO probe was used for few measurements (less than 6 hours) per day. As recommended by the manufacturer, the probe was dried and stored in a dry storage chamber with a silica-gel desiccant cartridge when not in use.

Calibration

Pre-calibrated LDO probe normally does not require calibration for normal operation, but calibration was ensured on daily basis by 100% water-saturated air.

B. Ammonia-Nitrogen Measurement

Ammonia-Nitrogen ion concentration was measured using the Hach **sension™** 2 ISE meter in combination with the 'Hach Combination Ammonia Electrode' (Figure 3-21).

Figure 3-21 ISE meter and Combination Electrode for Ammonia Concentration Measurement
(adopted from www.hach.com)⁵

The electrode measures ammonia gas or ammonium ions in aqueous solutions that have been converted to gas by the addition of a strong base. For most wastewater samples, 1 mL of 10 N NaOH is sufficient to increase the pH above 11. Dissolved ammonia in the sample solution diffuses through the membrane until the partial pressure of ammonia is the same on both sides of the membrane. The gas diffuses through the membrane and causes a pH change in the thin layer of electrolyte. The potential across the pH glass varies as a result of the pH change and the electrode measures the change in potential. The measured potential change is proportional to the ammonia concentration in the solution. This potential decreases by about 59.16 mV for every tenfold increase in ammonia concentration in the linear operating range at 25 °C. The electrode exhibits 95% response in one minute or less for ammonia concentrations above 0.68 mg/L NH₃ or 0.56 mg/L NH₃-N.

Reproducibility

Within the operating range of the electrode, reproducibility is independent of concentration. With proper calibration, measurements up to $\pm 2\%$ can be obtained.

Temperature Effects

A change in temperature causes the electrode's response to shift and change the slope. To eliminate this effect the calibration and measurement were performed at about the same temperature (room temperature).

⁵ Hach Company, 100 Dayton Avenue, Ames, Iowa 50010 Telephone: (515) 232-2533

Effect of Ionic Strength

Water vapor is a potential electrode interference which causes electrode drift. Adding a strong base to samples of low osmotic strength automatically adjusts them to the correct level. At high pH, ammonia solutions lose ammonia to the atmosphere, lowering the concentration. To eliminate such error, measurements were carried out as soon as possible after the solution becomes basic.

Interferences

Most gases do not interfere as they are converted to ionic form in basic solutions. Ionic species cannot cross the gas-permeable membrane and are not direct electrode interferences. Ammonia forms metal complexes with a number of metal ions. The use of strong bases inhibits the formation of some common metal complexes because of the high concentration of hydroxide ion. When hydroxide is present, only mercury will appreciably form a complex with ammonia. The use of synthetic wastewater also ensures the absence of these interfering metal ions.

Calibration

Ammonia-nitrogen concentration of samples from the RBC process was expected to fall from 200 mg/L in the influent to above 10 mg/L in the effluent. These concentrations are well within 0.5 to 14000 mg/L $\text{NH}_3\text{—N}$, the linear region for the ammonia electrode. Hence, a short direct two-point calibration method was adequate and was used.

C. Nitrate-Nitrogen Measurement

Nitrate ion concentration was measured using the Hach sension™ 2 ISE meter in combination with the Hach Platinum Series Nitrate Electrode (Figure 3-22). The combination electrode has the advantage of the simplicity of a single probe containing the reference

Figure 3-22 ISE meter and Combination Electrode for Nitrate Concentration Measurement
(adopted from www.hach.com)⁶

⁶ Hach Company, 100 Dayton Avenue, Ames, Iowa 50010 Telephone: (515) 232-2533

The nitrate electrode has a solvent-polymer membrane half-cell and a reference half-cell combined in one body. The membrane half-cell uses a low solubility ion exchanger in a polyvinyl chloride (PVC) matrix to detect nitrate ions. The double junction reference half-cell uses bridging ammonium sulfate reference electrolyte gel, and has a nonclogging free-diffusion junction. Reference electrolyte contacts the inner electrolyte. The reference half-cell uses a silver / silver chloride reference element in a sealed inner chamber, which establishes a fixed potential when in contact with the internal filling solution. The nitrate ion selective membrane undergoes ion exchange with nitrate in the sample, creating a potential across the membrane which varies with the amount of nitrate ion in the sample. This potential will decrease by about 58 mV for every tenfold increase in nitrate concentration in the linear operating range at 25 °C. Compared with colorimetric methods, the electrode method has several advantages. These are: no interference by sample color or turbidity; simple and rapid procedure; and comparable accuracy and precision.

Temperature Effects

The calibration curve slope changes by 0.33% for every 1 °C change in temperature. The change in the slope magnifies at low nitrate concentrations. To eliminate this effect the calibration and measurement were performed at about the same temperature (room temperature).

pH Effects

The nitrate electrode can be used over a pH range of 3 to 11. Normally, measurements are made at pH 3 to 4 to eliminate carbonate and bicarbonate interference and reduce organic acid interference. The Nitrate buffer solution adjusts the pH to between 3 and 4.

Ionic Strength Effects

The probe actually senses nitrate ion activity, not concentration. The activity is a product of the concentration and activity coefficient. The activity coefficient varies with ionic strength. Hence, to measure nitrate concentration, raising the ionic strength to a high value is the common method of ensuring the same ionic strength for all samples and standards. Ionic strength was controlled by adding Nitrate buffer solution in standards and water samples.

Interferences

No ion selective electrode is perfectly free of interferences from other similarly charged species. For the nitrate electrode, the major interferences include perchlorate, iodide, nitrite, bromide and chloride. Use of synthetic wastewater ensures the absence of these ions. The addition of Nitrate buffer solution eliminates most of these interferences in case ions are present.

Calibration

As nitrate concentration of the sample was expected anywhere between 0-10 mg/L and the linear region is between 4 mg/L to 14,000 mg/L NO₃—N a short direct two-point calibration method was not sufficient. A low-level multi-point calibration was used to cover a wider range of nitrate concentrations that has both linear and non-linear electrode response region. A multi-point calibration also ensures more accurate measurements than a two-point calibration.

D. COD Measurement

COD test samples were digested using Bioscience 163-486 COD reactor. Chemical Oxygen Demand was measured using Orbeco 975-MP portable water analysis system. (Figure 3-23).

COD Reactor was capable to hold up to 15 reagent tubes (16mm) for digestion at 150°C. The COD meter is an advanced microprocessor-controlled pre-calibrated photometer. It contains seven precision filters which are fixed inside, and is used for highest measuring sensitivity for various tests for water and wastewater. It can also be used as colorimeter or photometer, by calibrating for own special tests

COD Reactor

Photometer

Figure 3-23 COD Reactor and Photometer⁷ used for COD Test

Calibration

The COD reactor is factory set to maintain the correct temperature at 150°C. The 975-MP is factory pre-calibrated for 90 important water and wastewater tests, including COD tests ranging from very dilute to high strength. The instrument's memory for various calibrations can be accessed by entering a 'test code' for each test.

⁷ Orbeco Analytical Systems, Inc.

E. pH Measurement

The pH was measured using the Hanna HI 9025C pH meter in combination with the HI 1230B pH electrode (Figure 3-24). HI 9025C is a microprocessor-based pH meter in waterproof casing. HI 9025C measures both pH and temperature. HI 1230B pH electrode is a combination gel-filled double junction single probe containing the reference. The temperature probe can be used independently for temperature measurements, or it can be used in conjunction with the pH electrode to utilize the meter's automatic temperature compensation capability.

Temperature Effects

The temperature has an effect on the pH. The calibration buffer solutions are affected by temperature changes to a lesser degree than normal solutions. Temperature compensation for pH values can be achieved by using automatic built-in feature or manually using a chart. All pH values were corrected manually to 25 °C using the above mentioned chart (Appendix E - 1).

Figure 3-24 pH Meter with Electrode and Temperature Sensor⁸

Calibration

The pH meter has 5 memorized buffer values (4.01, 6.86, 7.01, 9.18 and 10.01), buffer recognition to avoid errors during calibration and automatic temperature compensation. The meter allows 1 or 2 point calibration using any of the 5 memorized buffers. As pH values of all samples were below 6.3, two-point calibration was carried out daily using 4.01 and 6.83 pH buffers.

3.11 Sampling & Test Methods

The process flow diagram (Figure 3-2) indicates the sampling points and the analytical parameters. Table 3-13 summarizes the sampling protocol and test methods used in this study. All grab samples

⁸ Hanna Instruments Canada Inc. 3156, Boul. Industriel, Laval, Québec (Canada) H7L 4P7 Toll Free 1-800-842-6629

were collected from the aeration chambers close to the entry point in stage 1 and from the RBC cradle at the tail-end of each stage. The collected samples were immediately filtered through the 1.5 µm Whatman filter paper in order to remove interfering suspended particles. The analyses of the filtered samples were conducted as per APHA et al. (1995) as soon as possible to circumvent the potential experimental error resulting from the biological activity in the sample during storage. The ambient temperature under which the tests were conducted ranged between 22 °C and 26 °C. During the course of continuous operation, parameters such as pH, temperature, DO, ammonia-nitrogen and nitrate-nitrogen were measured. The RBC performance parameters were recorded using the form shown in Appendix F - 5.

Table 3-13 Sampling and Analytical Protocol

Parameters	Sampling Points	Frequency	Sampling method	Sample Qty.	Sample Pretreatment	Analytical Instrument	Other Equipments used	Test Method
Dissolved Oxygen	F, A, 1 to 6	Twice a Day	Immersion	Bulk	None	Hach LDO meter	-	Luminescent dissolved oxygen (LDO) method D 888-05 *
pH	F, A, 1 to 6	Twice a Day	Immersion	Bulk	None	Hanna ISE meter	-	Electrometric Method 4500-H ⁺ B
Temperature	F, A, 1 to 6	Twice a Day	Immersion	Bulk	None	Built-in Thermistor (Hach LDO meter)	-	-
COD	F, A, 1 to 6	Daily	Grab	100 ml	Filtered using 1.5 µm Whatman filter	Orbeco 975-MP Photometer	'Bioscience' COD reactor	Closed Reflux Colorimetric Method 5220D
NH ₄ ⁺ -N	F, A, 1 to 6	Daily	Grab	100 ml	Filtered using 1.5 µm Whatman filter	Hach ISE meter with Ammonia Selective electrode	-	Ammonia-Selective Electrode method 4500-NH ₃ F
NO ₃ ⁻ -N	F, A, 1 to 6	Daily	Grab	100 ml	Filtered using 1.5 µm Whatman filter	Hach ISE meter with Nitrate Selective electrode	-	Nitrate- Electrode method 4500-NO ₃ ⁻ D

* Approved Standard Test Method D 888-05 – Interim Approval issued by ASTM on January 03, 2006

F- Feed Tank; A – Stream Leaving Aeration Tank; 1 to 6 – Streams leaving RBC Stages 1 to 6 respectively

4. RESULTS AND DISCUSSION

This chapter discusses the results from the studies conducted by the author in the 'Laboratory of Water Treatment Technologies'. As discussed earlier in Section 3.4, the study was carried out in two phases. In Phase-I, steady-state dissolved oxygen profiles were developed with clean water without any biomass developed in the RBC unit. Results of phase-I study are discussed in Section 4.1. In phase-II was assessed the performance of RBC for simultaneous removal of organics and nitrification. Also, the unit energy consumption by both the main drive and the aeration blower were studied. Experimental results are represented graphically in Appendix G. The performance of RBC unit based on the results of phase-II study is discussed in Section 4.3. As discussed earlier in Section 3.7, it took 45 days to fully develop biofilm on the media and to stabilize the operating conditions in the RBC unit. This occurred due to some problems with experimental setup and some operational difficulties. The details of progression of the seeding process and the problem faced in this process are discussed in Section 4.2. The energy consumption by RBC is discussed in Section 4.4.

4.1 Observations and Interpretation of the Experimental Results Obtained from Phase-I study – Oxygen Transfer in the Absence of Biomass in the RBC

This section is devoted to the discussion of the results from phase-I. In phase-I, steady-state dissolved oxygen profiles were developed for various combinations of the angular velocity and the rate of aeration. The experimental matrix used to conduct this study is presented in Table 3-6. Deaerated tap water was fed continuously to the RBC at the feed rate of 1 L/min ($\approx$ 1500 L/day). The RBC was operated at different combinations of angular velocities and aeration rates. Angular velocities were selected in the range of 2.5 RPM to 5 RPM and the rates of aeration were selected between 0 SCFM and 15 SCFM. The temperature of the influent deaerated water varied between 13.0°C and 17.6 °C. In a few cases, the temperature dropped to 12.1 °C or rose to 19.4 °C. This occurred for the initial 10 to 15 minutes during the start-up in the morning, after overnight stoppage. Dissolved oxygen levels of deaerated water varied between 1.2 and 2.2 mg/L all over phase-I.

Dissolved oxygen and temperature at all the sample points were measured every 15 minutes till steady state was reached. Steady-state was assumed when the last 3 readings for each sample point were within ± 0.1 mg/L range. The last set of readings was considered as steady-state DO concentrations. During the entire phase-I, the temperatures of the water in the RBC at various sampling points ranged

between 13.3 °C and 18.6 °C. To make comparable the DO profiles at various operating conditions, the measured DO concentrations were corrected to 25 °C. The temperature during phase-II varied around 25 °C. For this reason, 25 °C instead of 15 °C was selected to correct DO concentrations.

Figure 4-1 to Figure 4-4 show steady-state DO profiles for various combinations of angular velocity and rate of aeration. In Figure 4-2 to Figure 4-4, it is clearly visible that when deaerated water was aerated in the aeration chamber, it became supersaturated to 105% to 107% depending on the rate of aeration. This was achieved in the first 10 to 15 minutes once the aeration was started. Influent deaerated water fed was only 12 to 20 % saturated. This shows that without any biological oxygen demand, only the aeration chamber was more than capable of supplying the oxygen demand for deaerated clean water with the constant flow rate of 1500 L/day.

When this supersaturated water flowed through the RBC stages, DO concentration dropped by about 0.3 mg/L irrespective to the rate of aeration which was equivalent to 3 % of saturation concentration. It shows that the rotating discs were not capable of keeping the level of super-saturation achieved in the aeration chamber. However, the DO level in the RBC stages was still above the saturation concentration. Angular velocities did not show any significant effect on DO concentrations. DO concentrations in the last four stages of RBC were almost identical when wastewater in the aeration chamber was aerated.

Figure 4-1 shows steady-state DO profiles when the RBC is operated without any aeration in the aeration chamber. In this case the deaerated tap water was directly fed to the stage 1 of the RBC instead of feeding to the aeration chamber. This was done to avoid DO enrichment of the deaerated water in the aeration chamber due to longer residence time. The DO concentration of the water leaving the first stage was in the range of 90% to 98% saturation depending on the angular velocity. DO concentrations in stage 2 reached or passed saturation level. In the final stages, DO concentration was about 102% and remained almost invariable. This shows that for this particular flow rate of 1500 L/day of deaerated tap water, only 2 stages were enough to supply the oxygen demand.

4.1.1 Discussion of Oxygen Mass Transfer Coefficient in the RBC

Larger volumes of aeration chamber and stages of the RBC, and smaller liquid flow rate resulted in longer residence time. Residence time in the aeration chamber was 20.16 hours, and in the RBC stages ranged from 1.6 hours to 2.9 hours. The total residence time in the cradle was 11.44 hours. Hence, the aeration chamber and each individual stage of the RBC can be considered as a series of completely mixed flow reactor and equation (2-16) can be used to calculate oxygen mass transfer coefficient. The terms C_0 and C in equation (2-16) are replaced by $C_S - C_0$ and $C_S - C$ respectively.

Figure 4-1 Steady State Dissolved Oxygen Profiles for Various Angular Velocities with no Aeration in Aeration Chamber

Figure 4-2 Steady State Dissolved Oxygen Profiles for Various Angular Velocities at 5 SCFM Rate of Aeration in Aeration Chamber

Figure 4-3 Steady State Dissolved Oxygen Profiles for Various Angular Velocities at 10 SCFM Rate of Aeration in Aeration Chamber

Figure 4-4 Steady State Dissolved Oxygen Profiles for Various Angular Velocities at 15 SCFM Rate of Aeration in Aeration Chamber

When water in the aeration chamber was aerated, steady-state DO concentrations in the aeration chamber as well as in the RBC stages were always higher than saturation DO concentration. As the term $(C_s - C)$ in equation (2-16) was negative, it was not possible to determine the oxygen mass transfer coefficient. Similarly, determination of the oxygen mass transfer coefficient for stages other than stage 1 of the RBC was not possible. However, the same discs are used in the entire RBC, the oxygen mass transfer coefficient determined for stage 1 can be applied to all the stages of the RBC. Table 4-1 shows the calculated values of the mass transfer coefficients, K and K_L for RBC using experimental data for various angular velocities.

Table 4-1 Determination of Oxygen Mass Transfer Coefficients for Stage 1 of the RBC Based on Experimental Data

Volume of Stage 1 of the RBC	V	180	Liters
Total Surface Area of Discs in Stage 1	A	21.5	m ²
Volumetric Flow Rate	Q	1	L/min
O ₂ Saturation Concentration	C _s	8.09	mg/L

Angular Velocity (RPM)	Steady-State DO Concentration (mg/L)		$K \text{ (min}^{-1}\text{)}$ $= \left[\left(\frac{C_s - C_0}{C_s - C} \right) - 1 \right] * \left(\frac{Q}{V} \right)$	$K_L \text{ (m/min)}$ $= K \left(\frac{V}{A} \right)$	$K_L \times 10^6 \text{ (m/s)}$
	Influent (C ₀)	Stage 1 -at Exit (C)			
2.5	1.4	7.3	0.0415	0.000347	5.7894
3	1.5	7.5	0.0565	0.000473	7.8833
3.5	1.2	7.7	0.0926	0.000775	12.9199
4	1.1	7.7	0.0940	0.000787	13.1187
5	1.6	7.9	0.1842	0.001542	25.7038

Kim and Molof (1982), Ravetkar and Kale (1981) and Castilla Rodriguez (2001) models were selected to compare the experimental results. These models were suitable because they were developed by correlating experimental data for oxygen mass transfer without biofilm. Especially Castilla Rodriguez (2001) model was developed in similar conditions and for the same RBC which was used for this research.

Table 4-2 Comparison of Experimental Oxygen Mass Transfer Coefficient with the Values Calculated Based on Various Models

Diameter of Discs	D	0.88	m
Diameter of Discs *	D	34.646	inches
Disc Spacing *	d	21.5	inches
Molecular Diffusivity of Oxygen	D_m	2.1×10^{-9}	m^2/s

Angular Velocity ω (RPM)	$K \times 10^2 \text{ (min}^{-1}\text{)}$		$K_L \times 10^6 \text{ (m/s)}$		
	Present Study	Kim & Molof (1982)	Present Study	Castilla Rodriguez (2001)	Ravetkar and Kale (1981)
		$= 0.0011 \left(\frac{\omega^{1.5} D^{0.5}}{d} \right)^{0.732} *$		$= (2.49\omega + 1.45) \times 10^{-6}$	$\frac{K_L D}{D_m} = 1.59 \left(\frac{D^2 N}{D_m} \right)^{0.5}$
2.5	4.149	1.906	5.789	7.675	115.206
3	5.650	2.329	7.883	8.920	126.202
3.5	9.259	2.759	12.920	10.165	136.314
4	9.402	3.194	13.119	11.410	145.726
5	18.421	4.081	25.704	13.900	162.927

* Values for disc diameter, (D) and disc spacing (d) must be in inches

The model developed based on experimental data for the same RBC could be the reason for closed agreement of the experimental results with Castilla Rodriguez (2001) model. The slight difference in the results could be explained by the difference in the operational mode of the RBC. In 2001 it was operated the RBC in a batch mode, whereas this study was carried out in continuous operational mode. On other hand, the experimental results did not agree with Ravetkar and Kale (1981) model. They used very small flat acrylic discs of only 125 to 175 mm diameter compared to the 0.88 m diameter commercial discs with matrix used in the present study. Experimental results slightly agreed with Kim and Molof (1982) model. Those investigators carried out their study using flat discs with diameter ranging from 0.15 to 0.60 m.

According to Friedman et al (1979) oxygen mass transfer coefficients can be correlated to the angular velocity by the empirical equation:

$$\ln K_L = \frac{a}{\ln \omega} + b$$

(4-1)

Where

K_L = Gas-Liquid Oxygen mass transfer coefficient, m/s

ω = Angular Velocity of the Discs, RPM

a and b = Empirical Constants

Figure 4-5 Prediction of Empirical Constants for Friedman et al (1979) Model by Fitting to the Experimental Results Obtained in the Present Study

Figure 4-5 exhibits the experimental results of this research fitting to this model with coefficient of determination R^2 at 0.9241. Equation (4-1) can be rewritten as follows by replacing the empirical constants a and b by their values derived from statistical analysis (Figure 4-5).

$$\ln K_L = \frac{-3.0172}{\ln \omega} - 8.8757 \quad (4-2)$$

Figure 4-6 demonstrates the power correlation fitting experimental data more significantly, with the coefficient of determination, R^2 at 0.9675.

Figure 4-6 Prediction of Mathematical Model for Oxygen Mass Transfer by Fitting Power Correlation to the Experimental Results

Equation (4-3) can be considered a better mathematical model for correlating the oxygen mass transfer coefficient in this RBC and angular velocity of the media.

$$K_L = 0.8216 \times 10^{-6} (\omega)^{2.1047} \quad (4-3)$$

4.2 Observations and Problems during the Seeding Process

This section is intended to discuss the evolution of the seeding process and also the various process control and operational problems faced during this process until the proper biofilm on media and suspended biomass growth in the aeration chamber were established. As discussed earlier in Section 3.7, a thin biofilm was visible within three days of the operation. It became relatively thick in the first and second stages and was relatively thin in the final stages on the 8th day. That is, before the transition from the batch to the continuous mode. Figure 4-7 shows the development and progression of the biofilm in all the six stages of the RBC.

On the 8th day, the RBC was operated under continuous mode with a feed rate of 0.1 L/min. The angular velocity and the rate of aeration were 1.5 RPM and 5 SCFM respectively. The DO concentration in the aeration chamber was 7.75 mg/L. When this wastewater was entering into the stage 1 of the RBC, DO concentration was dropping to 1.24 mg/L and was recovering to 3.86 mg/L in stage 6 when flowed through the RBC cradle.

Up to Day 13, everything seemed to be in order. The feed rate, angular velocity and aeration rate were increased gradually to 0.810 L/min, 2.3 RPM and 7 SCFM respectively. The COD of the influent and effluent from the last stage were 5970 mg/L and 1648 mg/L respectively, which exhibited 72.4% COD removal. The DO concentration in the aeration chamber dropped to 0.48 mg/L. DO concentrations in the RBC stages ranged between 0.11 mg/L and 0.22 mg/L. This could be associated to the high organic load and intense bioactivity in the RBC. Furthermore, DO concentration in the feed tank dropped to 0.24 mg/L, which showed that significant bioactivity and bacterial growth were established in the feed tank.

From Day 14 onwards, the biofilm color started changing and on the 16th day, it became bright greenish yellow. The pH in feed tank dropped to 2.7. This was happened as a result of undesirable bioactivity due to longer residence time of synthetic wastewater in the feed tank. The pH was adjusted using 10 N NaOH solution. From Day 16 onwards the pH was adjusted to 7 - 8.5 using 10N NaOH.

From Day 17 the RBC started recovering and the biofilm color was changing back to normal. The feed rate, angular velocity and rate of aeration were 0.750 L/min, 3 RPM and 5 SCFM respectively. The pH in the RBC was between 6.39 and 7.08. DO in the last 3 stages were higher than 5 mg/L but the first 2 stages were still experiencing very low DO (< 0.2 mg/L). The DO concentration in the feed tank dropped to less than 0.2 mg/L. Bioactivity due to longer residence time of influent wastewater in the feed tank was causing low DO concentration as well as a significant drop in the pH. Large amount of biomass was settled in the feed tank, which made mandatory to flush it clean the feed tank frequently.

For about 2 weeks the same practices of adjusting the pH and flushing the feed tank before the experiments were continued. Variations of the pH resulted in severe loss of biofilm twice during this period. Also, once the synthetic wastewater in the feed tank became septic over the weekend. This required completely emptying and flush clean the feed tank. The wastewater in the RBC was also affected by receiving septic water. In order to recover quickly, about 60-70% of the water from the aeration chamber and about 80% water from the RBC cradle was drained and replaced by fresh synthetic wastewater.

After about 4 weeks of operation and facing difficulties to maintain the proper influent conditions, it was decided to provide a very short residence time to the synthetic wastewater in the feed tank. Hence, 160 L square section tank replaced the 5000 L feed tank (Figure 3-7, Figure 3-17). This allowed preparing synthetic wastewater continuously with less than 100 L of inventory at any time during the experiment. pH adjustment was discontinued and allowed microorganism in the bioreactor to adjust the suitable environment themselves. This solved the problem. In 2 weeks, stable biofilm on the media and suspended biomass in the aeration chamber were developed. The whole seeding and startup process took about 45 days.

Figure 4-7 Development and Progression of Biofilm on RBC media

4.3 Evaluation of the Experimental Results obtained in Phase-II – Performance of the RBC Treating High Strength Synthetic Wastewater

This section is intended to discuss the experimental results achieved in phase-II. The interpretation of the results is focused on the process parameters and chemical composition of the wastewater during phase-II. Performance related to the energy consumption is discussed in the following Section 4.4.

The collection of data in Phase-II started once the proper growth of the biofilm on rotating media and the suspended biomass in the aeration chamber were achieved. Over the entire phase-II the RBC unit was fed continuously with synthetic wastewater at the constant feed rate of 1 L/min (≈ 1500 L/day) which resulted into the consistent hydraulic loading of $0.017 \text{ m}^3/\text{m}^2 \text{ day}$. The composition of the synthetic wastewater (Table 3-10) was maintained consistent throughout the study, which resulted in the organic and ammonia-nitrogen loading to the RBC unit of $55.97 \text{ gBOD}/\text{m}^2\text{-day}$ and $3.444 \text{ gNH}_3\text{-N}/\text{m}^2\text{-day}$ respectively. Table 3-11 explains various operating and process parameters such as flow rate, residence times in the aeration chamber as well as in each stage of the RBC and loads to the RBC unit. The RBC was operated at various combinations of angular velocity and rate of aeration till the steady state was attained. The experimental matrix is shown in Table 3-7. Steady state was ensured by consistent DO concentrations and pH values at all the sampling points. Various operating and process parameters were measured and logged during the course of each run of the experiment, using the forms shown in Appendix F - 3 to Appendix F - 5. Steady state characteristics of the wastewater at different stages of the RBC unit for each run are represented graphically in Appendix G - 1 through Appendix G - 16.

4.3.1 Effect of Rate of Aeration and Angular Velocity on Dissolved Oxygen Concentrations

A. Effect of Rate of Aeration on DO Concentration in the Aeration Chamber

The relationship between the rate of aeration and the DO concentration in the aeration chamber is shown in Figure 4-8. The DO concentration in the aeration chamber ranged from 5.5 mg/L to 6.1 mg/L when wastewater was aerated at 15 SCFM . It is well known that DO concentrations between 2 mg/L to 4 mg/L are favorable for aerobic processes (Tchobanoglous et al., 2003). The DO concentrations in the aeration chamber at 15 SCFM were far more than what is favorable. On other hand, aeration at 15 SCFM created a great deal of turbulence in the aeration chamber which may not be required for efficient or optimum performance. Higher concentration of the DO represents the opportunity for optimizing the performance by turning down the blower and saving the energy. This possibility for optimization is also depending on further treatment and the performance in the cradle.

Figure 4-8 Correlation of DO Concentration in the Aeration Chamber with Rate of Aeration

When the aeration rate reduced to 12 SCFM, the DO concentration in the aeration chamber dropped to 3.7 mg/L, which is well within the favorable range but on a higher side.

With 10 SCFM aeration rate, the DO concentrations were ranging between 2 mg/L and 2.5 mg/L, which is on the lower side of the favorable range. This shows an opportunity to optimize the rate of aeration between 10 SCFM and 12 SCFM.

When the rate of aeration was reduced to 8 SCFM and 5 SCFM, the DO concentrations in the aeration chamber dropped to 0.7 and 0.2 mg/L respectively. These resulted into DO limiting condition and loss of substrate removal performance. At 5 SCFM, the COD removal and ammonia removal performance dropped to 34% and 17% respectively from their corresponding percentage removal performance of 58% and 36% respectively for 10 SCFM. Furthermore, insufficient mechanical mixing at these aeration rates resulted into large accumulation of the biomass in the aeration chamber and in turn viscous mixed liquor. When this mixed liquor entered and flowed through the RBC, eventually resulted into the viscous wastewater in the whole RBC unit. This probably affected the RBC performance too. This means that a rate of aeration lower than 10 SCFM is not recommendable. On other hand, the mechanical mixing and physical characteristics of the wastewater in the aeration chamber at 10 SCFM and 12 SCFM were observed well compared to other aeration rates studied.

B. Effect of the Angular Velocity on DO Concentration in the Rotating Disc Zone

After leaving the aeration chamber, the wastewater flowed through the RBC stages. The DO concentrations in the rotating disc zone depend on the oxygen mass transfer capacity of the media, which in turn depends on the angular velocity. Furthermore, these DO concentrations are also depending on the loading to the stages. The loading to stage 1 of the RBC depends on the performance of the aeration chamber, which in turn depend on rate of aeration in the aeration chamber. Hence, the relationships of DO concentrations with angular velocities are presented for various rates of aeration (Figure 4-9 and Figure 4-10).

Figure 4-9 Correlation of DO Concentration in Stage 1 with Angular Velocity

Figure 4-9 shows the relationship of DO concentrations in stage 1 with angular velocities, plotted separately for each rate of aeration. The DO concentration in stage 1, at 15 SCFM rate of aeration in combination with 2.5 RPM and 3 RPM were well within the favorable range. DO concentrations for the combinations of 4 RPM with 12 SCFM and 10 SCFM aeration rates were also well within the favorable range.

Similarly, Figure 4-10 shows the relationship of the angular velocity with DO concentrations of the effluent leaving the RBC cradle. Each curve represents the different aeration rate in the aeration chamber. Angular velocities between 3 RPM and 4 RPM, and aeration rate between 8 SCFM to 10 SCFM were observed as favorable operating parameters for optimum performance.

Figure 4-10 Correlation of DO Concentration in Stage 6 with Angular Velocity

In the previous discussion in Section 4.3.1 (A), 10 SCFM and 12 SCFM were identified as favorable rates of aeration compared to other rates. Combining these observations with the previous observations shows that the opportunity for optimum performance lays around 10 SCFM of the aeration rate and 4 RPM of the angular velocity.

4.3.2 COD and Ammonia Removal Performance

This section discusses COD and ammonia removal performance of the aeration chamber and the rotating disc zone separately as well as combined in the RBC unit. The removal performance of the aeration chamber is discussed in Section 4.3.2 (A). Section 4.3.2 (B) discusses the COD and ammonia removal performance of the rotating disc zone as a separate unit. The COD and ammonia removal performance of the RBC unit as a whole, is discussed in Section 4.3.2 (C)

A. COD and Ammonia-Nitrogen Removal in the Aeration Chamber

This experiment was carried out at constant hydraulic, organic and ammonia-nitrogen loadings of $0.017 \text{ m}^3/\text{m}^2\text{-day}$, $86.108 \text{ gCOD}/\text{m}^2\text{-day}$ ($\approx 55.97 \text{ gBOD}/\text{m}^2\text{-day}$), and $3.444 \text{ g NH}_3\text{-N}/\text{m}^2\text{-day}$ respectively and at different rate of aeration in the aeration chamber. However, variations in the influent synthetic wastewater and probable errors in the analysis caused the COD loading to vary between 83.5 and $90.6 \text{ gCOD}/\text{m}^2\text{-day}$, and ammonia-nitrogen loading between 3.2 and $3.9 \text{ g}/\text{m}^2\text{-day}$. The only variable operating parameter for the aeration chamber which is correlated to the removal performance is the

rate of aeration. That means the COD and ammonia removal performance in the aeration chamber depend on the rate of aeration only. Figure 4-11 shows the relationships of % COD and ammonia-nitrogen removals with the rate of aeration in the aeration chamber. Values of coefficient of determination confirm the significant correlation between removal rates and rate of aeration in the aeration chamber. A significant amount of COD removal in the aeration chamber was observed, which ranged from 34% to 67% for rate of aeration ranged from 5 SCFM to 15 SCFM respectively. An average 65% COD removal rate can be estimated for 15 SCFM rate of aeration from the regression fit (Figure 4-11). DO concentration in the aeration chamber was always observed higher than the favourable range of 2 mg/L to 4 mg/L. Removal rate slowly declined to 61%, 58% and 55% when the rate of aeration reduced to 12 SCFM, 10 SCFM and 8 SCFM respectively. When the rate of aeration decreased from 8 SCFM to 5 SCFM the COD removal rate was sharply dropped to an average of 34%. This means a rate of aeration lower than 8 SCFM is not optimum. This shows the opportunity for energy saving with reasonably good performance lies within the range of 8 SCFM and 12 SCFM.

Figure 4-11 Correlation of % COD and NH₃-N Removals in the Aeration Chamber with Rate of Aeration

Curves for % COD removal and % NH₃-N removal with the rate of aeration showed almost similar trends (Figure 4-11). This means that the oxidation of the ammonia is in proportion to the consumption of the organic matter. This could be associated to the consumption of ammonia by heterotrophic bacteria and absence of autotrophic nitrifying bacteria. An average 41% ammonia-nitrogen removal was observed at 15 SCFM rate of aeration. Like the COD removal rate, the NH₃-N removal rate slowly

declined to 38%, 34% and 32% when rate of aeration decreased to 12 SCFM, 10 SCFM and 8 SCFM. The removal rate sharply declined to 17% when the rate of aeration reduced to 5 SCFM. It means reducing only 3 SCFM of air supply can halve the efficiency. This reconfirms the range of 8 SCFM to 12 SCFM as the potential range for optimizing the RBC performance and achieving minimum unit energy consumption.

B. COD and Ammonia-Nitrogen Removal in the RBC Cradle

Unlike the loadings of COD and $\text{NH}_3\text{-N}$ to the aeration chamber, loadings to the RBC cradle were not consistent. These loadings depended on the performance of the aeration chamber. When influent aerated in the aeration chamber, COD and $\text{NH}_3\text{-N}$ loadings to stage 1 of the RBC varied between 28.278 and 60.103 $\text{gCOD/m}^2\text{-day}$, and between 2.053 to 3.048 $\text{gNH}_3\text{-N/m}^2\text{-day}$. These ranges were extended to 91.998 $\text{gCOD/m}^2\text{-day}$ and 3.780 $\text{gNH}_3\text{-N/m}^2\text{-day}$ when the RBC was operated without any aeration. In this case influent was fed directly to stage 1 of the RBC in order to avoid biological activity and septic conditions which can arise as a result of 20 hours residence time in the aeration chamber.

Figure 4-12 Correlation of % COD Removal with COD Loading in Rotating Disc Zone

Relationships of % COD and ammonia-nitrogen removals with their respective loadings are presented in Figure 4-12 and Figure 4-13 respectively. Removal rates in the RBC trough depend not only on their respective loadings but also on the angular velocity of the media. Each data point on the graphs represents different angular velocity of the media. Hence, it was very difficult to develop any correlation between rate of loadings and their removal rates. The scattered distributions of the data points and the very small values for the coefficients of determination confirm that there is no correlation.

Figure 4-13 Correlation of % $\text{NH}_3\text{-N}$ Removal with $\text{NH}_3\text{-N}$ Loading in Rotating Disc Zone

C. Overall Performance of RBC unit for COD and Ammonia-Nitrogen Removal

In the previous sections the aeration chamber and the rotating disc zone were evaluated individually for their COD and $\text{NH}_3\text{-N}$ removal performance. In this section the aeration chamber and the rotating disc zone are combined as an RBC unit for evaluation. The overall COD removal performance of the RBC unit combining the performance of aeration chamber and rotating disc zone is presented in Figure 4-14.

When influent wastewater was aerated at a rate of 10 SCFM and above and the angular velocity was set at 4 and 5 RPM, higher than 93% of the COD removal efficiencies were achieved. Maximum COD removal rate of 95.8% was observed at 15 SCFM and 5 RPM. When the aeration rate was turned down to 12 SCFM and 10 SCFM keeping the same 5 RPM angular velocity, the COD removal rates decreased to 94.7% and 93.8% respectively. This means by reducing 2% of COD removal efficiency, significant energy saving can be achieved. The reduction in the rate of aeration below 10 SCFM showed a significant loss in performance, this can be observed from the nature of the curve irrespective of the angular velocity of the rotating media (see Figure 4-14).

When the angular velocity of the media was reduced from 5 RPM to 4 RPM while keeping the rate of aeration constant at 15 SCFM, the COD removal rate was reduced by just 0.58%. Similarly at 10 SCFM on reducing angular velocity from 5 to 4 RPM, the removal rate was dropped by only 0.48%. This means that 4 RPM can be suitably selected as an operating speed compared to 5 RPM by trading only about 0.5% of overall COD removal efficiency.

Figure 4-14 COD Removal Performance of the RBC Unit (Combined Aeration Chamber and Rotating Media)

Keeping the angular velocity constant at 4 RPM and reducing the rate of aeration from 10 SCFM to 8 SCFM, 5 SCFM and no aeration, the COD removal rate sharply dropped to 87%, 71.2% and 63.8% respectively. A significant steep decrease of the curve between 10 SCFM and 5 SCFM can be observed in Figure 4-14. This occurred because of insufficient mechanical mixing and accumulation of

biomass in the aeration chamber. As a result the wastewater became viscous and the suspended biomass became very difficult to settle. No provision to allow biomass settle in the aeration chamber and no mechanism to extract the settled sludge from the aeration chamber made this situation worse. The removal efficiency of the rotating media was also affected when this liquid eventually entered and flowed through the RBC cradle.

Similar kind of phenomenon was observed even for the combination of 3 RPM and 5 SCFM. This phenomenon was observed only when the liquid in the aeration chamber was aerated.

When the RBC was operated without any aeration and the wastewater fed directly to the RBC cradle, the removal efficiency still dropped, but not at the pace at which the aeration rate reduced from 10 SCFM to 5 SCFM. The curve between 5 SCFM and 0 SCFM is relatively less steep than that of the curve between 10 and 5 SCFM (Figure 4-14). Furthermore, the wastewater in the RBC cradle was very clear and containing easily settleable sloughed biofilm. This phenomenon observed even with the angular velocity as low as 2.5 RPM. That is, the performance deteriorated only because of the accumulation of biomass in the aeration chamber. A sludge settling zone and sludge withdrawal arrangement can improve the overall performance of the RBC unit and the unit energy consumption.

When the rotating disc zone only was operated without any aeration in the aeration chamber, the removal efficiency dropped and ranged between 34% and 68.8% depending on the angular velocity. In contrast, the removal efficiency ranged between 77.3% and 95.8% when aeration chamber was aerated at the rate of 15 SCFM. This significant improvement achieved with aeration can potentially reduce the footprint of the treatment unit.

Figure 4-15 shows the removal of ammonia-nitrogen with various combinations of angular velocities and rates of aeration. The profiles of the curves for ammonia-nitrogen removal rate showed almost the same profiles as COD removal rate curves (see Figure 4-14 and Figure 4-15). The maximum ammonia-nitrogen removal rate of 68.08% was observed at 5 RPM and 15 SCFM. This removal rate slowly decreased to 62.63% when the rate of aeration decreased to 10 SCFM, and sharply dropped to 45.2% when the RBC was operated without any aeration and the wastewater directly fed to the rotating disc zone. This 17% to 23% gain in the removal rate by aeration can justify the additional aeration system.

The $\text{NH}_3\text{-N}$ removal rate was negligibly reduced by only 0.04%, when the angular velocity was reduced from 5 RPM to 4 RPM keeping the rate of aeration constant at 15 SCFM. This shows that 4 RPM is a better speed to achieve energy conservation objectives without sacrificing much on removal performance. When the angular velocity was kept constant at 4 RPM and the rate of aeration reduced

from 15 SCFM to 12 SCFM and 10 SCFM, the removal rate slowly declined from 68.04% to 66.1% and 59.9% respectively.

Figure 4-15 Ammonia-Nitrogen Removal Performance of the RBC Unit (Combined Aeration Chamber and Rotating Media)

Figure 4-16 Removal Performances of Conventional and Modified RBCs

Further reduction in the rate of aeration to 8 SCFM and 5 SCFM caused the removal rate sharply drop to 50.5% and 45.1% respectively. In contrast, the NH₃-N removal dropped very little to 42.9% when aeration was discontinued. This can be clearly observed from the nature of the curve (Figure 4-15). The drop in the removal rate between 5 SCFM and 0 SCFM is not as steep as the drop between 10 SCFM and 5 SCFM. This can be associated to the reduced performance of the aeration chamber at 5 SCFM and 8 SCFM. With no aeration, 21.2% to 45.2% NH₃-N removal was observed compared to 45.6% to 68.1% for 15 SCFM rate of aeration. The 23% to 24% improvement in ammonia removal efficiency can justify the addition of the aeration system to the RBC. This can effectively reduce the footprint and achieve the same level of treatment. The removal performances of conventional and modified RBCs are summarized in Figure 4-16.

4.3.3 Nitrification in the RBC Unit

Figure 4-17 and Figure 4-18 show the relation between ammonia-nitrogen converted to nitrate-nitrogen in the aeration chamber and in the rotating discs zone, with the rate of aeration and angular velocity respectively. Nitrate-nitrogen concentration in the influent wastewater was around 0.5 mg/L. This was resulted due to the availability of nitrate in the tap water. No nitrate was added in the preparation of synthetic wastewater. Maximum 0.932 mg/L of nitrate-nitrogen concentration was observed in the wastewater exiting the aeration chamber which corresponded to 0.23% conversion of ammonia-nitrogen to nitrate. In some cases the nitrate concentration in the aeration chamber was as low as 0.5 mg/L which corresponded to a conversion as low as 0.03%. This shows negligible to almost no nitrification in the aeration chamber. This little nitrification could be associated to the existence of some nitrifying bacteria by virtue of longer solids residence time in the aeration chamber.

Figure 4-17 % of Ammonia-Nitrogen Converted to Nitrate-Nitrogen in the Aeration Chamber

Figure 4-18 % of Ammonia-Nitrogen Converted to Nitrate-Nitrogen in the Rotating Discs Zone

Figure 4-18 shows the amount of nitrification achieved in the RBC cradle. Maximum 3.47 mg/L nitrate-nitrogen concentration was observed in the effluent from the RBC cradle, which corresponded to 2.21 % conversion to nitrate-nitrogen from the ammonia-nitrogen loaded to the RBC. This shows very little nitrification in the RBC cradle.

It is well known that autotrophic nitrifying bacteria cannot grow in the presence of organic matter in high concentration (Kissel et al., 1984; Wanner and Gujer, 1984, 1986; Furumai and Rittmann, 1992; Rittmann and Manem, 1992; Okabe *et al.*, 1995; Wik and Breitholtz 1996). In order to establish themselves nitrifiers required BOD to reduced upto 30 mg/L and they can outperform if BOD drops to 10 mg/L. A minimum of 205 mg/L of COD ($\approx$ 135 mg/L of BOD₅) was observed in the effluent. Hence, nitrification was not expected. Little nitrification achieved in the RBC can be interpreted as the presence of nitrifiers deep into the thick biofilm. Lower organic matter concentration in the interior of the biofilm created the suitable environment for their existence and in turn the little conversion of ammonia to nitrate.

4.4 Comparison of the Power Drawn by the Main Drive and the Blower with and without Biomass

As discussed earlier in chapter 3, the power consumption was studied with and without biomass developed in the RBC unit. This section is intended to discuss the power consumption by the main drive and the aeration blower. Section 4.4.1 discusses the comparison of the energy consumption by the main drive when the RBC operated with and without biofilm. Section 4.4.2 discusses the energy consumption by the aeration blower to aerate tap water as well as to aerate wastewater containing suspended biomass. The theoretical power requirements for the main drive and the aeration blower are calculated and compared with the actual power consumption. Section 4.4.3 discusses the unit energy consumption for different combinations of the angular velocity and the rate of aeration. Also, in this section the optimum angular velocity and rate of aeration to achieve minimum unit energy consumption is discussed.

Actual energy consumptions by the main drive and the aeration blower were measured using two separate KWH meters placed before the variable frequency drive. The variable frequency drives were used to control the speed of the main drive motor and the aeration blower. Energy consumptions in KWH and time were measured. The power drawn (KW) by the motor and the blower were calculated by dividing the energy consumption (KWH) by the time (hours) for which the motor or the blower were operated.

4.4.1 Energy Consumption by the Main Drive

Figure 4-19 shows a comparison of the power (KW) drawn by the main drive motor with the RBC operated with and without biofilm on the media. A correlation between the angular velocity and the actual power drawn over the entire operating range of the main drive is presented in Appendix D - 1. This correlation was developed with no biofilm on the rotating media. On this chart the power drawn by the main drive motor and the angular velocities are plotted against the motor speed which is

represented in terms of % frequency. Energy consumption data for a specific range of angular velocities from this correlation were re-plotted against angular velocity (Figure 4-19). In Figure 4-19 the dotted curve represents the regression fit to these power consumption data.

Figure 4-19 Power Drawn by the Main Drive Motor with and without Biofilm on the Rotating Media

The solid line curve represents the regression fit to the actual power consumption data for the main drive when the RBC was operated with high-strength synthetic wastewater and the media was covered with biofilm. Each data point on this curve represents the actual power drawn by the main drive during the experimental runs in phase-II.

The resultant regression equations (see Figure 4-19) were used to calculate the power drawn by the main drive motor at various angular velocities. The calculated power consumptions by the main drive motor with and without biofilm on rotating media are tabulated and compared in Table 4-3.

The wired power is the actual power drawn by the main drive motor which includes all mechanical and electrical energy losses. The additional 6.18 to 11.61 watts power were drawn by the main drive motor when the discs were covered with the biofilm. This was equivalent to 18.97% to 24.10% in addition to the power drawn by the main drive without biofilm. The additional power drawn could be correlated to the additional torque required to rotate the biofilm developed on the media.

Table 4-3 Comparison of Wired Power Drawn by the Main Drive Motor with and without Established Biofilm

Angular Velocity (RPM)	Wired Power – Power Drawn by Main Drive Motor (Watts)		Additional Power Drawn when Disc Media covered with biofilm Compared to without Biofilm	
	Without Biofilm	With Biofilm	Watt	%
2.5	32.584	38.765	6.182	18.97%
3	36.977	45.786	8.810	23.82%
4	48.208	59.827	11.619	24.10%
5	62.701	73.868	11.167	17.81%

As discussed earlier the power consumption was measured by the KWH meter connected before the variable frequency drive. This means various electrical and mechanical losses are included in the measured power. These losses are associated with the efficiencies of the main drive motor, worm gear and chain drive. The theoretical power absorbed by the rotating media can be calculated by multiplying the wired power by the overall efficiency of the main drive system. Table 4-4 shows the estimation of the overall efficiency and the theoretical power absorbed by the rotating media.

Table 4-4 Estimation of Power Absorbed by the Rotating Media, from the Actual Power and the Estimated Efficiencies of the Main Drive Components

Rated Efficiency of the Main Drive motor, $\eta_{\text{main drive motor (rated)}}$ 68 %

Rated Efficiency of the Worm Gear, $\eta_{\text{main drive gear}}$ 40 %

Rated Efficiency of the Chain Drive, $\eta_{\text{chain drive}}$ 99 %

Wired Power Drawn by the Main Drive the Motor – without biofilm (Watts)	% Speed of motor	Overall Efficiency #	Power Absorbed by media, P_w (Watts) (= Wired power * Overall Efficiency)
		= % Speed of motor * $\eta_{\text{main drive motor (rated)}}$ * $\eta_{\text{main drive gear}}$ * $\eta_{\text{chain drive}}$	
32.5835	44.43 %	11.96 %	3.8983
36.9767	52.85 %	14.23 %	5.2623
48.2084	66.79 %	17.99 %	8.6704
62.7005	79.76 %	21.48 %	13.4667

Efficiency of main drive motor varies with speed - Efficiencies of Main Drive are estimated by multiplying the % Frequency of VFD and Rated Efficiency of Main Drive

Fujie et al. (1983) developed an empirical equation to calculate the theoretical power requirement by the discs without any biofilm. The empirical constant used in this equation can be determined by using

the estimated values of the power absorbed by the rotating media at different angular velocities. The estimation of the empirical constant from the power absorbed by the media is represented in Table 4-5.

Table 4-5 Estimation of Empirical Constant λ_1 from the Power Absorbed by the Media

Diameter of Discs, D (m)	0.88
Disc Spacing, d (m)	1.20×10^{-2}
Periphery of Discs, P (m)	2.765
Surface Area, A (m ²)	87.1
Density of Clean Water at 25°C, ρ (kg/m ³)	997
Dynamics Viscosity of Clean Water at 25°C, μ (Ns/m ²)	0.89×10^{-3}

Angular Velocity, N (RPM)	Tip Velocity, (m/s)	Reynolds Number, Re $= \frac{NDd\rho}{\mu}$	$N^2 D^2$	P_w/A	λ_1 $= \frac{P_w/A}{N^2 D^2}$
2.5	0.115	492.90	4.84	4.48×10^{-5}	9.25×10^{-6}
3.0	0.138	591.48	6.97	6.04×10^{-5}	8.67×10^{-6}
4.0	0.184	788.64	12.39	9.95×10^{-5}	8.03×10^{-6}
5.0	0.230	985.80	19.36	15.46×10^{-5}	7.99×10^{-6}

Laminar flow condition can be correctly assumed in the RBC cradle as the tip velocities were less than 0.5 m/s (Fujie et al., 1983) for the range of the angular velocities used. Hence, equations (2-26) to (2-30) (2-31) were used to calculate the constant λ_1 . Table 4-5 shows the estimation of the empirical constant λ_1 . The calculated values of λ_1 are varying within a very short range of -0.83σ to $+1.28\sigma$. This variation could be correlated to the experimental errors and the assumption made for the efficiencies of the main drive components. The empirical constant λ_1 , could be more reliably estimated if the losses involved could be eliminated from the calculation. This could be attained by employing a torque meter to measure energy absorbed by the rotating media.

Table 4-6 Comparison of Experimentally Determined Value of Empirical Constant λ_1 , With the Values Available in the Literature

	$\lambda_1 \times 10^5$ (kW.min ² /m ⁴)	Disc Diameter (m)
Experimental	0.799-0.925	0.88
Sekine et al. (1979)	1.10-1.75	0.5
Antonie (1976)	0.86	3.0
USEPA (1971)	0.39	1.75

The experimentally determined values closely matched the values determined by Sekine et al. (1979) and Antonie (1976) (see Table 4-6). The difference in the results could be explained by the difference in the flow conditions, and the size and the geometry of the discs.

4.4.2 Energy Consumption by the Aeration Blower

Figure 4-20 shows a comparison of the power consumed by the blower for aeration of tap water and synthetic wastewater containing suspended biomass. A correlation between the rate of aeration and the actual power drawn over the entire capacity of the blower is presented in Appendix D - 2. This correlation was developed in the absence of any suspended biomass. On this chart the power drawn by the blower and the rate of aeration are plotted against the blower speed which is represented in terms of % frequency. The energy consumption data for a specific range of aeration rates from this correlation were re-plotted against angular velocity (Figure 4-20). In Figure 4-20 dotted curve represents the regression fit to these power consumption data which can be mathematically presented by the following equation (4-1).

$$P_{W_{blower}} = 200.77 e^{0.0563 F_{air}} \quad (4-1)$$

Where,

$P_{W_{blower}}$ Actual wired power drawn by aeration blower, kW

F_{air} Requisite power, SCFM

The solid curve represents the regression fit to the actual power consumption data when high strength synthetic wastewater containing suspended biomass was aerated. This correlation can be mathematically presented by the equation (4-2). Each data point on this curve represents the actual power drawn by the blower during experimental runs in phase-II.

$$P_{W_{blower}} = 201.89 e^{0.0564 F_{air}} \quad (4-2)$$

Where,

$P_{W_{blower}}$ Actual wired power drawn by aeration blower, kW

F_{air} Requisite power, SCFM

Figure 4-20 Comparison of Power Drawn by the Blower for Aeration of Clear Tap Water and Synthetic Wastewater Along With Suspended Biomass

The resultant regression equations (see Figure 4-20) were used to calculate the wired power drawn by the aeration blower at various rates of aeration. Calculated power consumptions by the blower to aerate tap water and synthetic wastewater containing suspended biomass are tabulated and compared in Table 4-7.

Table 4-7 Comparison of Wired Power for the Aeration Blower with and without Suspended Biomass

Rate of Aeration (SCFM)	Theoretical Power Requirement, P_{adb} (Watts) [□]	Wired Power – Power Drawn by Blower Motor (Watts)*		Additional Power Drawn when Wastewater with Suspended Biomass Aerated		Overall Efficiency $= P_{adb} / P_{w_{blower}}$
		Clear Water	Wastewater with Suspended Biomass	Watt	%	
5	23.93	266.044	267.662	1.618	0.61%	8.94%
8	41.05	314.996	317.007	2.011	0.64%	12.95%
10	53.60	352.539	354.860	2.321	0.66%	15.10%
12	67.74	394.556	397.233	2.677	0.68%	17.05%
15	91.91	467.155	470.466	3.311	0.71%	19.54%

□ Calculated based on equation (2-35)

* Calculated based on the regression equations (Figure 4-20)

Both regression curves were almost identical and matching each other (Figure 4-20). This means that the actual wire power to aerate clear water and wastewater with suspended biomass were almost identical. Aeration blower drew only 1.6 to 3.3 KW extra to aerate wastewater with suspended biomass, which was only 0.61% to 0.71% in addition to what the blower consumed to aerate clear water. This could be explained by the identical specific gravity and physical characteristics of wastewater and clear water, which caused almost similar pressure drop in the aeration system.

The calculated theoretical power requirement and the overall efficiency of the blower are tabulated in Table 4-7 and are also represented by Figure 4-20. The theoretical power requirement for the blower was calculated using equation (2-35). The overall efficiency of the blower was calculated dividing the theoretical power by the actual power. The overall efficiency of the blower includes the efficiencies of the blower and the blower motor.

For an aeration rate of 15 SCFM the variable frequency drive was set at about 92 % (≈ 55 Hz). The calculated overall efficiency at 15 SCFM was 19.54% (Table 4-7). This calculated efficiency was closely matched the efficiency reported by the supplier (see the performance curve Appendix C - 4). The performance curve shows the efficiencies of the blower for frequencies 55 Hz and above, hence it was not possible to compare efficiencies for other aeration rates.

4.4.3 Unit Energy Consumption in the RBC

In this section unit energy consumption for various combinations of angular velocity and rate of aeration are compared and discussed. As mentioned earlier unit energy consumption is defined as the amount of energy consumed (KWH) per unit of oxygen demand removed. In the literature most of the studies for energy consumption in RBC considered unit energy consumption as KWH energy consumed per kg of BOD₅ removed or stabilized. Antonie (1976) suggested UOD (Ultimate oxygen Demand) instead of BOD₅ to calculate unit energy consumption.

In the present study wastewater samples were analyzed for COD instead of BOD₅. Wherever BOD₅ is referred, the values are calculated assuming BOD₅ as 65% of COD. In the subsequent discussion the unit of energy consumption is referred to both UOD and COD. Equation (2-37) was used to calculate the Ultimate Oxygen Demand (UOD).

A. Unit Energy Consumption with no aeration (Only Main Drive)

The RBC used in the present study, was modified by incorporating a new aeration system. Most of the energy consumption study for the RBC refers to the conventional RBC that is in the absence of additional aeration. In this study, the RBC can be considered as a conventional RBC, when operated without any aeration. The influent synthetic wastewater was fed directly to stage 1 of the RBC. Hence,

this energy consumption data could be directly compared to the unit energy consumption reported in the literature.

Table 4-8 Comparison of the Experimentally Determined Unit Energy Consumption in This Study with the Value from Literature

	Benschine and Fok (2002)	Present Study
Type of Wastewater	Domestic	High Strength Synthetic
Flow Rate (m³/day)	6813	1.5
Influent BOD (mg/L)	172	3060-3472
Effluent BOD (mg/L)	12	1117-2078
% BOD Removal	93 %	34%-67%
Unit Energy Consumption (KWH/Kg BODr)	1.1	0.52 – 0.68

The unit energy consumption observed ranged between 0.34 and 0.44 KWH/kg CODr, when the RBC operated with no aeration in the aeration chamber. In terms of BOD₅ removal, this unit energy consumption is equivalent to 0.52-0.68 KWH/kg BODr. This seemed to be very low compared to 1.1 KWH/kg BODr reported by Benschine and Fok (2002) and others. This vast difference in the results could be related to the difference in the capacity of the treatment units, organic and hydraulic loads, geometry of the disc assembly and removal performances. Table 4-8 compares various parameters from these studies.

Lower unit energy consumption in the present study corresponds to the lower removal performance. COD removal rate was reported between 34% and 66.8% compared to 93 % in the literature. The most important parameter is geometry of the discs. The substrate used in this study was table sugar which was comparatively easily biodegradable than the domestic waste in their study. It is obvious that the terminal treatment (prior to discharge) demands more energy than the gross removal. The RBC in present study reduced the BOD₅ from an average of 3266 mg/L to about 1600 mg/L which was very easy to remove compared to BOD₅ from 172 mg/L to 12 mg/L in their study. Antonie (1976) showed that the energy consumption to remove the organic load from 50% to 95% could be up to 5 folds than what is required to remove initial 50%.

B. Unit Energy Consumption with Aeration

The relationship between the unit energy consumption (kWH/kg CODr) and various combinations of angular velocity and the rate of aeration is presented in Figure 4-21. Figure 4-23 represents the same data in 3 dimensions, excluding the unit energy consumption by the RBC without aeration.

Figure 4-21 Unit Energy Consumption (kWh/kg of CODr) for Various Combinations of Aeration Rate and Angular Velocity

Figure 4-22 Correlation of Unit Energy Consumption (KWH/kg CODr) and Rate Aeration at 4RPM Angular Velocity

The maximum unit energy consumption observed during the entire study was 2.03 kWh/kg CODr, for the combination of 2.5 RPM and 15 SCFM. The unit energy consumption for 15 SCFM in combination with various angular velocities ranged from 1.75 to 2.03 kWh/kg CODr. For 15 SCFM, a minimum unit

energy consumption was observed at 4 RPM. The highest unit energy consumptions were recorded with 15 SCFM compared to other rates. The aeration blower contributed 87 to 92% of total power drawn by the RBC unit. However, the blower contributed to remove 69 to 85% of the total COD removed in the RBC unit. This shows an inefficient use of energy. Higher DO concentrations in the aeration chamber at 15 SCFM were the evidence of wastage of energy which were not utilized to remove the oxygen demand. This energy can be saved by turning down the blower.

Figure 4-23 3-Dimensional Representation of Unit Energy Consumption (kWh/kg of CODr) for Various Combinations of Aeration Rate and Angular Velocity (excluding unit energy consumption with no aeration)

This is exactly what was confirmed when the blower's flow was turned down to 10 SCFM. The unit energy consumption reduced significantly and ranged from 1.31 to 1.67 kWh/kg CODr. At 10 SCFM, the minimum unit energy consumption (1.31 kWh/ kg CODr) was observed at 4 RPM. This was observed earlier for 15 SCFM. This shows that the optimum angular velocity is closed to 4 RPM. The aeration blower contributed 83 to 91% of total power drawn by the RBC unit for 62 to 76% of total COD removal. This shows that aeration blowers are the major contributor of the energy consumption.

Thus, the disproportionate energy consumption by the blower relative to the COD removal justifies both the optimization of the operation of the systems like the one here studied and the research undertaken herein. This optimization can be empirically achieved by maintaining DO level in the aeration chamber within the favourable range 2-4 mg/L.

At 5 SCFM, the aeration blower contributed 83 to 87% of the total power drawn by the RBC unit and removed 47 to 55% of total COD removal. The unit energy consumption was observed to be between 1.43 and 1.66 kWh/kg CODr. The unit energy consumption was increased by 0.12 to 0.26 kWh/kg CODr compared to the unit energy consumption for 10 SCFM, for their corresponding angular velocities. This shows an opportunity to optimize unit energy consumption around 10 SCFM. Combining this observation with the observation made earlier, it can be interpreted that the optimum point lies around 4 RPM and 10 SCFM. For this reason two more aeration rates, 8 and 12 SCFM, were selected to study the performance and unit energy consumption. The unit energy consumption at 8 SCFM and 12 SCFM were found at 1.34 and 1.54 kWh/kg CODr, which were respectively 0.03 and 0.2 kWh/kg CODr higher than the unit energy consumption for the combination of 4 RPM and 10 SCFM. This shows that local optimum is situated between 8 and 12 SCFM.

The relationship between the unit energy consumption (kWh/kg UODr) and various combinations of angular velocities and the rates of aeration is presented in Figure 4-24. The same data are represented in 3 dimensions, excluding the data for energy consumption without aeration in Figure 4-26. UOD includes the oxygen demand for ammonia-nitrogen removal in addition to the demand for organic matter removal.

Figure 4-24 Unit Energy Consumption (KWH/kg UODr) for Various Combinations of Aeration Rate and Angular Velocity

Figure 4-25 Correlation of Unit Energy Consumption (KWH/kg UODr) and Rate Aeration at 4 RPM Angular Velocity

Figure 4-26 3-Dimensional Representation of Unit Energy Consumption (kWh/kg of UODr) for Various Combinations of Aeration Rate and Angular Velocity (excluding unit energy consumption with no aeration)

Identical shapes of these curves with the curves for the energy consumption per unit COD removal (kWH/kg CODr) reconfirms the absence or minor impact of the autotrophic nitrifying bacteria, and showed that most of the ammonia-nitrate was consumed as a result of organic matter removal.

C. Optimum Operating Parameters for Minimum Unit Energy Consumption

Local optima for the unit energy consumptions with respect to COD, BOD and UOD removals, were determined using the 'Table curve 3d by Jandel Scientific' software. Appendix H shows a detailed statistical analysis. Optimum combinations of angular velocity and rate of aeration to achieve minimum energy consumption are summarized in Table 4-9.

Table 4-9 Optimum Combination of Angular Velocity and Rate of Aeration to Achieve Minimum Unit Energy Consumption

Angular Velocity (RPM)	Rate of Aeration (SCFM)	Unit Energy Consumption
3.66	8.13	1.3068 (KWH/kg CODr)
3.66	8.13	2.0105 (KWH/kg BODr)
3.66	8.13	1.1969 (KWH/kg UODr)

Appendix H shows the regression fit equations for each case, which can be used as mathematical models to calculate the unit energy consumptions for a given angular velocity and rate of aeration. All regression equations showed good fit with coefficient of determinations, $R^2 > 0.99$.

4.5 Error Analysis

No measurement is perfectly accurate or exact. Many instrumental, physical and human limitations cause measurements to deviate from the 'true' values of the quantities being measured. These deviations are called "experimental uncertainties," or more commonly "errors".

Various experimental errors were expected while collecting the experimental results. Experimental errors indicate unavoidable variations rather than faults or mistakes. Experimental errors are associated with variations in the environmental conditions, unavoidable variations in operating and process control parameters, imperfections in collecting samples, errors associated with measuring and test equipments, variations in the standards and reference solutions, personnel skills and other aspects. Experimental errors refer to the fluctuation or discrepancy in replicate observations from one experiment to another.

Errors related to the personnel skills are very difficult to quantify. Care was exercised in collecting and preparing the samples, in preparation of standards and reference solutions. All test instruments used

for analytical purposes were maintained and stored by following the suppliers' recommendations. The majority of the errors which can be quantified were errors involved with accuracy of measuring and testing equipment. Process and analytical parameters measured are listed in Table 3-12 and Table 3-13 along with the measuring and test equipment used. All the instruments used for the analytical purpose are recognised internationally as precise instruments. The accuracy of the instruments used for the analytical and process control purpose are listed in Table 4-10.

Table 4-10 Errors Associated With Measuring and Test Equipments

Measuring and Test Equipment	Accuracy
Water Flow Totalizer	0.5 %
Water Flow meter	6% of full scale
Air Flow meter (Rotameter)	2% of full scale
Tachometer	0.006 % of Reading
kWH Meter	0.2 %
COD Reactor	$\pm 2^{\circ}\text{C}$
pH meter	$\pm 0.01 \text{ pH}; \pm 0.5^{\circ}\text{C} (@ 20^{\circ}\text{C})$
LDO meter	DO: $\pm 0.1 \text{ mg/L}$ (at $<8 \text{ mg/L}$); $\pm 0.2 \text{ mg/L}$ (at $>8 \text{ mg/L}$) Temperature: $\pm 0.3^{\circ}\text{C}$ Pressure: $\pm 2\%$ of reading
ISE meter (Ammonia and Nitrate Analysis)	Milivolt: $\pm 0.2 \text{ mV}$ or $\pm 0.05\%$ of reading, whichever is greater Temperature: $\pm 1^{\circ}\text{C}$

In addition to the aforementioned errors, there were some operational and analytical errors involved as listed below.

- The concentrated synthetic wastewater was dosed into the feed tank and diluted continuously using tap water. The variations in the flow rates of tap water and concentrated solution generated up to $\pm 10\%$ errors in the compositions of synthetic wastewater (see Appendix G).
- Steady state was assumed based on DO and pH profiles. The error caused by this assumption can not be quantified.
- The results were analysed based on a single set of data for each run rather than on an average.

5. CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

The rising cost of energy, clean water crisis and availability of land are demanding highly compact and energy efficient wastewater technologies with excellent removal performance. RBCs with added aeration systems can meet these requirements. An added aeration system improves the performance but demands more energy than conventional RBCs. Hence, the relevance to optimize the operating parameters. The optimization of the operating conditions leads to major energy savings by sacrificing little on removal performance. To achieve this objective a major modification of the previous experimental unit located in the Laboratory of Water Treatment Technologies was carried out. That included

- Enlargement of the aeration chamber,
- Replacement of the coarse bubble diffusers by fine bubble diffusers,
- Replacement of the central air supply by regenerative air blower, and
- Replacement of the mono-block motor by a three phase motor with variable frequency drive

This thesis has focused on three main areas of research these are: (a) Oxygen Mass Transfer; (b) Performance of the modified RBC; and (c) Energy Consumption

A. Oxygen Mass Transfer in RBC Operated with Tap Water

- A DO concentration of 105-107% (oversaturation) was achieved after 10-15 minutes of operation with fine bubble diffusers aerating deaerated tap water.
- Full DO saturation was also achieved in the first two stages of the conventional RBC.

B. Performance of the the RBC Treating High Strength Synthetic Wastewater

To study the removal performance and optimization, the modified RBC was fed with synthetic high-strength wastewater containing 5000 mg/L of COD and 200 mg/L of $\text{NH}_3\text{-N}$ at a constant flow rate of

1500 L/day. The RBC was operated with various combinations of angular velocities between 2.5 and 5 RPM and rates of aeration between 0 and 15 SCFM. These are the most relevant results:

- Depending on the rate of aeration, 34-67% of COD and 17-45% of $\text{NH}_3\text{-N}$ removal were observed in the aeration chamber.
- Depending on the angular velocity, the conventional RBC used herein achieved 34-67% COD removal and 21-45% $\text{NH}_3\text{-N}$ removal.
- Depending on the rates of aeration and angular velocities 42-96% COD removal and 40-68% $\text{NH}_3\text{-N}$ removal were achieved in the modified RBC.
- When the RBC was operated with aeration, it showed up to 43% improvement in COD removal and 24% in $\text{NH}_3\text{-N}$ removal. This means added aeration can potentially reduce the footprint of the treatment unit.
- Depending on both the rates of aeration and the angular velocities, 1.09-3.05 mg/L $\text{NO}_3^-\text{-N}$ concentration was observed at the exit of modified RBC.
- The treatment with the RBC only, produced 0.93-2.33 mg/L of $\text{NO}_3^-\text{-N}$ depending on the angular velocity. Insignificant amount of nitrate concentration shows the absence of biofilm dominated by nitrifying bacteria. Little nitrification achieved, could be attributed to the presence of nitrifiers deep into the thick biofilm.
- Rates of aeration below 8 SCFM create inadequate mechanical mixing, excessive biomass accumulation and very low DO levels in the aeration chamber.

C. Energy Consumption Optimization

The present study indicates that the hybrid bioreactor, based on fixed film and suspended growth can be optimized by analyzing the removal efficiency and energy consumption of each process and selecting the operating conditions for minimum energy consumption per unit oxygen demand removed.

- Depending on the rates of aeration and angular velocities, the unit energy consumption ranged from 1.2 to 1.89 KWH/kg UODr in the modified RBC, and 0.31-0.39 KWH/kg UODr in the conventional RBC. However, only 34-67% of COD and 21-45% of $\text{NH}_3\text{-N}$ removal was achieved in conventional RBC.
- Depending on the rate of aeration, the blower consumed 83-92% of the total energy and removed 34-67% of COD and 21-45% of $\text{NH}_3\text{-N}$.
- A maximum removal of 96% COD and 67% $\text{NH}_3\text{-N}$ were achieved at 5 RPM and 15 SCFM. However, the minimum unit energy consumption of 1.20 KWH/kg UODr was achieved at 4 RPM and 10 SCFM.

- Statistical analysis of the experimental data showed that 3.66 RPM angular velocity and 8.13 SCFM aeration rate are the optimum parameters to achieve a minimum unit energy consumption of 1.1969 KWH/kg UODr.

5.2 Recommendations for Future Research

- The same study should be repeated after making the necessary changes to remove sludge from the aeration chamber at various frequencies. This would allow evaluating the effect of sludge residence time on the reactor performance.
- This study was carried out at fixed hydraulic and organic loadings. Using the same statistical tools used in this work we could expand the ranges to include various loadings, different C/N ratios, different geometries of the discs and RBC flow configurations. These could be utilized to develop more generalized mathematical models.

Preparing synthetic wastewater and preventing bioactivity due to microbial contamination in the 5000 L feed tank was a challenge during this experimental work. Furthermore, sludge accumulated during the operation interfered with the performance of the unit. Therefore, the following improvements are proposed:

- A smaller feed tank with better control system and real time synthetic wastewater preparation can be considered as a substitute of the 5000 L feed tank. Alternatively, concentrated solution and tap water could be directly and independently fed to the RBC.
- A drain valve or pump must be provided to remove sludge frequently from the aeration chamber and secondary clarifier. This would prevent anaerobic conditions and foul odor.
- There is no arrangement to empty out the RBC cradle. A small transfer pump or sludge pump can resolve this problem.
- Online measurement and the addition of a data logging system would make simpler to identify steady state conditions.

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Appendix A. Equipment Specifications

Appendix A - 1 Specifications of Process Equipment, Measuring Instruments and Test Instruments used in the Present Study

A. Vacuum Stripper

Column:

Material of construction		Transparent PVC
Diameter	m	1.03
Total Height	m	2.43
Wall Thickness	cm	1
No. of Sections		4
Section		Height (m)
Demister		0.18
Packed Bed Section (Upper)		0.85
Packed Bed Section (Lower)		0.70
Reservoir		0.70
Total		2.43

Packed Column:

No of Packed Beds		2
Packing Material		Tri-Packs®
	Upper	Lower
Packed Bed Height (m)	0.5	0.6
Packing Nominal Diameter (cm)	2.5 & 5	5

Tri-Packs® Packing:

		2.5 cm Ø	5 cm Ø
Packing Factor	1/m	91.86	52.49
Void Space	%	90	93.5
Geometric Surface Area	m ² /m ³	279	157
Bulk Density	Kg/m ³	99.40	67.34

Tripacks® – Jaeger Products Inc., 1611 Peach leaf St., Huston, TX 77039
Ph (281) 449-9500; 1-800-678-0345 Web site: www.jaeger.com

B. Feed Tanks and Dosage Tank

Material of construction	Type	Size	Capacity
High Density Polyethylene	Vertical cylindrical	1.5m Φ x 3m H	5000 Liter
High Density Polyethylene	Vertical Square (2 Tanks)	46 cm x 45 cm x 77 cm	160 Liters (each)

C. Rotating Biological Contactor Characteristics

Treatment Capacity	300 L/day (with municipal wastewater)	
	Unit	Quantity
No. of Stages	-	6
Angular Velocity	RPM	0-6
Disc Submergence	%	40

Discs:

Disc Diameter		m	0.88
Disc Periphery		m	2.7646
Space Between Discs		cm	1.2
Disc Surface Area		m ² / disc	1.262
Stage No.	Number of Discs	Surface Area (m ²)	Volume (m ³)
1	17	21.5	0.180
2	10	12.6	0.100
3	10	12.6	0.100
4	10	12.6	0.100
5	12	15.2	0.120
6	10	12.6	0.115
Total	69	87.1	0.715

Chambers:

	Volume (m ³)
Aeration Chamber	1.260
Clarification Tank	0.610
Total	1.870

D. Vacuum Pump (Oil Ring Vacuum Pump)**Pump**

Manufacturer		Precision Pump Co.	
Model		D 150	
Free Air Displacement	Ultimate Vacuum	Efficiency (@ 1µm Hg)	Oil Capacity
150 LPM (5.3 CFM)	0.1 µm Hg	60 %	2.37 L

Motor

Manufacturer			General Electric		
Model			5KC38RN4x		
HP	RPM	Hz	Phase	Voltage	Amp
¾	1725	60	1	115/230	11.6/5.8

E. Transfer Pump (Progressive Cavity Pump)**Pump**

Manufacturer		Eagle Pump and Compressor Ltd.	
Model		EP67-CSQM	
Capacity (@ 1750 RPM, 70 °F)			Maximum Speed (for Water)
53 GPM (@ 0 PSI)	34 GPM (@ 35 PSI)	2000 RPM	

Motor

Manufacturer			US Electric Motors, W Germany		
Model			E315-TMP		
HP	RPM	Hz	Phase	Voltage	Amp
1½	1725	60	1	115/208-230	16.9/7.0

'Varitrac' reduction gear

Manufacturer		US Electric Motors, W Germany	
Model / Type / Frame		E153 Z06 33296 / TD / 2	
Speed Range		2160/360 (For 1800 Motor RPM)	

F. Main Drive**Motor**

Manufacturer				Lesson Electric Corporation (USA)		
Model				C4T17DH3E		
HP	RPM	Hz	Phase	Voltage	F.L. Amp	Efficiency
½	1725	60	3	208-230/460	1.9-2.0/1.0	68 %

Reduction gear

Manufacturer			David Brown (Textron power Transmission)		
Model			A028060 WRN-B		
Nominal Ratio		Output Torque (mechanical)		Efficiency	
60:1		110 lb-in		40%	

Chain Drive

Nominal Ratio	Number of teeth
4.9091:1	54 (Motor side) / 11 (RBC side)

G. Regenerative Blower

Manufacturer	FPZ Inc., Italy
Model	SCL V3

Blower

Maximum airflow	Maximum Continuous Pressure
49 CFM	64 InWG

Motor (Induction motor)

HP	RPM	Hz	Phase	Voltage	FL Amp
¾	3500	60	3	208-230/460	2.9-2.6/1.3

H. Feed Pump (Progressive Cavity Pump)**Pump**

Manufacturer		Continental Pump Co.	
Model		CP33-SSOMT	
Capacity (@ 1750 RPM, 70 °F)			Maximum Speed (for Water)
9.4 GPM (@ 0 PSI)	4.2 GPM (@ 50 PSI)	2000 RPM	

Motor

Manufacturer			US Electric Motors, W Germany		
Model			E512-TMP		
HP	RPM	Hz	Phase	Voltage	Amp
½	1725	60	1	115/208-230	9.6/4.8

'Varitrac' reduction gear

Manufacturer		US Electric Motors, W Germany	
Model / Type / Frame		C4T17DH3E / TD / 1	
Speed Range		2160/360 (For 1800 Motor RPM)	

I. Fine Bubble Diffusers

Manufacturer			Environmental Dynamics Inc.	
Model			Flexair™ 9" Disc diffuser (Std)	
Peak air flow	Design air flow	Design DWP	Diffuser Diameter	Active surface area per diffuser
7 SCFM	0.5-5.0 SCFM	8-12 inWG	10.9"	0.4 ft ² (380.9 cm ²)

J. Variable Frequency Drive (Penta-Drive AC Motor Speed Control)

Manufacturer		KB Electronics Inc.	
Model / Enclosure		KBAC-24D / NEMA 4X IP-65	
Input		115 V; Single phase AC; 16/10 Amp (max)	
Output		0-230 V ; 3 ph AC; 3.6 RMS Amp/Phase (max, cont.)	
Maximum Rating		1 HP / 0.75 kW	

K. Metering Pump

Manufacturer		Blue-White Industries		
Model		C-6250HV		
Capacity (@ 0 PSI)	Maximum Pressure	Voltage	Frequency	Power
32.9 GPH	5 PSIG	115 V	60 Hz	45 W

L. Water Flow Totalizer

Manufacturer		Actaris		
Model		Mulitmag (Multi-jet Water meter)		
Flow Rate	Pressure (Max)	Temperature (Max)	Accuracy	
1/8 to 20 GPM	150 PSIG	50 °C	0.5 %	

M. Water Flow meter

Manufacturer		King Instrument Co.		
Model		7330-3-6-1-0-42W		
Flow Range	Maximum Pressure	Maximum Temperature	Accuracy	
0.1-1 GPM	150 PSIG	54°C	6% of full scale	

N. Air Flow meter (Rotameter)

Manufacturer		Dwyer Instrument Inc., Indiana, USA		
Model		VFC-121-EC		
Flow Range	Maximum Pressure	Maximum Temperature	Accuracy	
0.25-25 SCFM	690kPa (100 PSIG)	49°C	2% of full scale	

O. Pressure Gauge

Manufacturer		Weiss Instruments Inc.		
Model	Type	Range	Graduation	
DG25P3 1/2" LM	Low Pressure Diaphragm Gauge	0-5 PSI	0.1 PSI	

P. Tachometer

Manufacturer		Check-Line	
Model		DT-207	
Range	Sensor	Display Update Time	Accuracy
Up to 99,999 RPM	Laser Diode	1 Sec	0.006 % of Reading

Q. kWH Meter

Manufacturer		P3 International	
Model		P4400	
Voltage	Amp (Max)	Power (Max)	Accuracy
115 V AC	15 A	1875 VA	0.2 %

R. COD Reactor

Manufacturer		Bioscience Inc.		
Capacity	Voltage	Phase	Hz	Watts
1-15 tube heater	120 Volt AC	1	50/60	100

S. Photometer (for COD Test)

Manufacturer	Orbeco Analytical Systems Inc.
Model	975-MP
Wavelength Selection	420, 490, 528, 535, 565, 608, 640 nm
Temperature range	(Operating): 0-40 °C (Storage) -10-+70 °C
Light Source	Pre-focused tungsten lamp
Detector	Silicon photodiode
Calibration (Zero and standardization set)	Automatic by microprocessor

T. pH meter with electrode

Manufacturer	Hanna Instruments			
Model	HI 9025C			
pH Electrode	HI 1230B combination gel-filled double junction pH electrode			
Temperature probe	HI 7669/2W			
Calibration	Automatic with 5 memorized std. buffers (4.01, 6.86, 7.01, 9.18, 10.01)			
Temperature Compensation	Automatic or manual from 0-100 °C			
Range	Resolution	Accuracy (@ 20 °C)	Offset calibration	Slope Calibration
0 -14 pH; 0-100 °C	0.01 pH; 0.1 °C	± 0.01 pH; ± 0.5 °C	± 1 pH	From 70 to 108%

U. LDO meter with electrode

Manufacturer	Hach Company
Model	LDO™ HQ10
Temperature Sensor	30K Ohm Thermistor
Compensation	Automatic; 0 to 50°C ; 400 to 1100 mBar ; 0 to 70 ppt (Salinity)
Minimum Flow Rate	Negligible, <5 cm/minute
Measuring Range	DO: 0.01-20 mg/L (ppm); 0-200% saturation Temperature: 0 to 50° C Pressure: 400 to 1100 mBar
Resolution	DO: 0.01 mg/L (ppm) 0.01% saturation Temperature: 0.1 °C Pressure: 1 mBar
Accuracy	DO: ± 0.1 mg/L (at <8 mg/L); ± 0.2 mg/L (at >8 mg/L) Temperature: ± 0.3°C Pressure: ± 2% of reading
Sensitivity	± 0.05% of Span
Detection Limit	0.01 mg/L (ppm)
Response Time at 20° C	To 95% of measured value, <30 seconds
Interferences	None (at concentrations typically found in DW and WW) Species evaluated: H ₂ S, pH, Na ⁺ , Al ⁺³ , Fe ⁺² , Fe ⁺³ , SO ₄ ⁻² , Cl ⁻ , Cl ₂ , ClO ₂ , Oil
Environmental Conditions	Operation: 0 - 50° C; 0 - 95% non-condensing relative humidity Storage: -20 to +60°C; 0 to 95% non-condensing relative humidity
D.O. Calibration Methods	No Calibration Req. (< 6hr/day Use) / Saturated Air Calibration

V. ISE meter

Manufacturer	Hach Company
Model	sension2
Measuring Range	ISE Mode: 0 to 19900 mg/L Milivolt Mode: -2000 to 2000 mV Temperature: -10 to 110° C
Resolution	ISE Mode: 0.1/0.01/0.001 (selectable) Milivolt Mode: 0.1 mV Temperature: 0.1 °C
Accuracy	Milivolt Mode: ± 0.2 mV or $\pm 0.05\%$ of reading, whichever is greater Temperature: $\pm 1^\circ\text{C}$
Instrument Drift	< 40 $\mu\text{V}/^\circ\text{C}$
Environmental Conditions	Operation: 0 - 50° C at 85% non-condensing relative humidity

W. Ammonia Electrode

Manufacturer	Hach Company
Model	51927-00
Electrode Type	Combination electrode
Ammonia Membrane Modules	Replaceable membrane modules
Electrode Storage	Dry / Ammonia Storage Solution.
Electrode Resistance	<500 Megohm
Concentration Range	0.06 to 17,000 mg/L NH_3 ; 0.05 to 14,000 mg/L $\text{NH}_3\text{-N}$
Temperature Range	Operating: 0 - 50 °C Storage: -40 to 60 °C
pH Range	> pH 11; Ionic Strength Adjustor raises pH above 11
Slope	-57 ± 3 mV/decade in linear concentration range
Response Time	95% response in three minute(s) or less in linear range

X. Nitrate Electrode

Manufacturer	Hach Company
Model	51920 (Platinum Series)
Electrode Type	Combination electrode
Reference Electrolyte	Ammonium Sulfate Reference Electrolyte
Minimum Sample Volume	25.0 mL
Range	0.05 mg/L (3.5×10^{-6} N) to 14,000 mg/L (1 N) NO_3^- -N
Linear Region	4 mg/L to 14,000 mg/L NO_3^- -N
Slope	-58 ± 3 mV/decade
Sample pH Range	3 to 11 (ISA should bring to 3.6)
Temperature Range	(All standards and samples should be at the same temperature.)
	Routine: 10-30 °C; Occasional: 10-50 °C; Storage: -40 to 50 °C
Response Time	10 to 60 seconds in linear range

Appendix B. Blower Sizing

Appendix B - 1 Estimation of Air Requirement and Sizing of Air Blower

Oxygen Demand Loading		
Flow Rate	L/day	1500
BOD ₅ Loading	mg/L	3250
	Kg/day	4.875
Ammonia – N Loading	mg/L	200
	Kg/day	0.3
Theoretical O₂ Requirement		
% of BOD and Ammonia Removal Req.	%	100
O ₂ requirement for BOD	mg/mg BOD	1
	Kg/day	4.875
O ₂ requirement for Ammonia – N	mg/mg NH ₃ -N	4.3
	Kg/day	1.29
Total Theoretical O ₂ Requirement	kg/day	6.165
Assumed Data **		
Alpha		0.60
Beta		0.95
Theta		1.024
% oxygen transfer, SOTE*	%	10%
AOR/SOR		0.45
Air Requirement Calculation		
Oxygen Requirement (Calculated)	kg/day	137.0
Oxygen in Air**	wt %	23%
Air Requirement	kg/day	595.652
Air density**	kg/m ³	1.2
Air flow rate	m ³ /min	0.34471
Conversion Factor	m ³ /SCFM	2.83E-02
Air Requirement	SCFM	12.173
Number of Diffuser Required		
Suggested Air flow rate per diffuser***	SCFM	3 (Operating Range 0.5-5)
Minimum No. of diffuser Required		5
No. of diffusers required to maintain symmetry		6
Sizing of Blower		
Liquid head above the diffuser	Ft	3
Pressure Drop (Diffuser) *	Ft	1
Aeration System Pressure Drop **	Ft	1
Total Head Required for Blower	Ft	5
Minimum Air flowrate required at 5 ft head	SCFM	15

* Assumed using supplier's performance curves. (Refer Appendix C)

** Assumed based on supplier's calculation and/or Tchobanoglous et al., 2003

*** Environmental Dynamics Inc. 5601 Paris Rd., Columbia, MO 65202, Tel: 573-474-9456, www.diffuserexpress.com

Calculation:**Theoretical O₂ Requirement (for BOD₅ Removal)**

$$\begin{aligned}
 &= (3250 \text{ g/m}^3 * 1.5 \text{ m}^3/\text{day}) / (1000 \text{ g/kg}) \\
 &= 4.875 \text{ Kg/day}
 \end{aligned}$$

Theoretical O₂ Requirement (for NH₃-N Removal)

$$\begin{aligned}
 &= (200 \text{ g/m}^3 * 1.5 \text{ m}^3/\text{day}) / (1000 \text{ g/kg}) \\
 &= 1.29 \text{ Kg/day}
 \end{aligned}$$

Total Theoretical O₂ Requirement

$$\begin{aligned}
 &= \text{Theoretical O}_2 \text{ Requirement (for BOD}_5 \text{ Removal+ for NH}_3\text{-N Removal)} \\
 &= 6.165 \text{ Kg/day}
 \end{aligned}$$

Actual O₂ Requirement

$$\begin{aligned}
 &= \text{Theoretical O}_2 \text{ Requirement} / [(\text{oxygen transfer, SOTE}) * (\text{AOR/SOR})] \\
 &= (6.165 \text{ Kg/day}) / (0.1 * 0.45) \\
 &= 137 \text{ Kg/day}
 \end{aligned}$$

Actual Air Requirement

$$\begin{aligned}
 &= \text{Actual O}_2 \text{ Requirement} / (\% \text{ of O}_2 \text{ in Air}) \\
 &= (137 \text{ Kg/day}) / 0.23 \\
 &= 595.652 \text{ Kg/day}
 \end{aligned}$$

Required Air Flow Rate

$$\begin{aligned}
 &= \text{Actual Air Requirement} / \text{Air Density} \\
 &= [(595.652 \text{ Kg/day}) / (1.2 \text{ Kg/m}^3)] / (1440 \text{ min /day}) \\
 &= 0.34471 \text{ m}^3 / \text{min} \\
 &= 12.173 \text{ SCFM} \quad (\because 1 \text{ SCFM} = 2.83 \times 10^{-2} \text{ m}^3 / \text{min})
 \end{aligned}$$

Suggested Air Flow Rate per Diffuser = 3 SCFM

Minimum Number of Diffuser Required

$$\begin{aligned}
 &= 12.173 \text{ SCFM} / (3 \text{ SCFM/diffuser}) \\
 &= 4.06 \approx 5 \text{ diffusers}
 \end{aligned}$$

Number of Diffuser Required to Maintain Symmetry

$$= 6 \text{ diffusers}$$

Appendix C. Performance Curves

EDI FlexAir™ 9" Disc Diffuser (Standard Membrane) General Oxygen Transfer Efficiency (OTE) Curves

Appendix C - 1 Fine bubble membrane diffuser Performance Curves

(Environmental Dynamics Inc. 5601 Paris Rd., Columbia, MO 65202, Tel: 573-474-9456, Fax: 573-474-6988 website: www.diffuserexpress.com)

Appendix C - 2 Air Blower (SCL V3) Performance Curve (Capacity v/s Frequency)

Appendix C - 3 Air Blower (SCL V3) Performance Curve (Absorbed Power v/s Frequency)

Appendix C - 4 Air Blower (SCL V3) Performance Curve (Efficiency v/s Frequency)

Appendix D. Calibration Curves

Appendix D - 1 Rotating Biological Contactor Main Drive - Correlation of Frequency with Angular Velocity, Amperage and Power

Air Blower Calibration (% Frequency - Speed - Power Correlation)

Appendix D - 2 Aeration Blower - Correlation of Frequency with Airflow Rate, Static Pressure, Amperage and Power

Appendix E. Data Used to Correct measured value for process temperature

Appendix E - 1 Chart for Manual Temperature Correction of pH Values

(adopted from Instruction Manual- HI 9025C Portable Waterproof pH Meter)
(www.hannainst.com)

TEMP °C	pH VALUES				
	4.01	6.86	7.01	9.18	10.01
0	4.00	6.98	7.13	9.46	10.32
5	4.00	6.95	7.10	9.39	10.24
10	4.00	6.92	7.07	9.33	10.18
15	4.00	6.90	7.04	9.27	10.12
20	4.00	6.88	7.03	9.22	10.06
25	4.01	6.86	7.01	9.18	10.01
30	4.02	6.85	7.00	9.14	9.96
35	4.03	6.84	6.99	9.10	9.92
40	4.04	6.84	6.98	9.07	9.88
45	4.05	6.83	6.98	9.04	9.85
50	4.06	6.83	6.98	9.01	9.82
55	4.07	6.84	6.98	8.99	9.79
60	4.09	6.84	6.98	8.97	9.77
65	4.11	6.85	6.99	8.95	9.76
70	4.12	6.85	6.99	8.93	9.75

Saturated Dissolved Oxygen Concentration in Water at 172 m Altitude

Appendix E - 2 Oxygen Solubility in Clean Water at Various Atmospheric Temperatures at 172 m Altitude

Appendix F. Forms Used to Record Experimental Results

Energy Consumption without Biomass (Main Drive - RBC)

[illegible]

**Appendix F - 1 Format Used to Log Observations for Energy Consumption Without Biomass
While Calibrating Main Drive VFD**

Energy Consumption without Biomass (Aeration Blower)

[illegible]

**Appendix F - 2 Format Used to Log Observations for Energy Consumption Without Biomass
While Calibrating Aeration Blower VFD**

Log Sheet for DO Profiles in RBC

[illegible]

Appendix F - 3 Format Used to Log Observations for Dissolved Oxygen Profiles in Phase-I Study

[illegible]

Appendix F - 4 Format Used to Log Observations for Operating Parameters and Energy Consumption Data in Phase-II Study

RBC Performance

Date		Sampling Point							
		1	A	1	2	3	4	5	6
	Collected At (Time)								
	Temp (°C)								
	DO (mg/L)								
	pH								
	COD (mg/L)								
	NH ₄ ⁺ - N (mg/L)								
	NO ₃ ⁻ - N (mg/L)								
	Collected At (Time)								
	Temp (°C)								
	DO (mg/L)								
	pH								
	COD (mg/L)								
	NH ₄ ⁺ - N (mg/L)								
	NO ₃ ⁻ - N (mg/L)								
	Collected At (Time)								
	Temp (°C)								
	DO (mg/L)								
	pH								
	COD (mg/L)								
	NH ₄ ⁺ - N (mg/L)								
	NO ₃ ⁻ - N (mg/L)								
	Collected At (Time)								
	Temp (°C)								
	DO (mg/L)								
	pH								
	COD (mg/L)								
	NH ₄ ⁺ - N (mg/L)								
	NO ₃ ⁻ - N (mg/L)								

Appendix F - 5 Format Used to Log Observations of RBC performance

Appendix G. Experimental Results Phase-II

Appendix G - 1 Performance of RBC with Angular Velocity 2.5 RPM and No Air to Aeration Chamber

Appendix G - 2 Performance of RBC with Angular Velocity 3.0 RPM and No Air to Aeration Chamber

Appendix G - 3 Performance of RBC with Angular Velocity 4.0 RPM and No Air to Aeration Chamber

Appendix G - 4 Performance of RBC with Angular Velocity 5.0 RPM and No Air to Aeration Chamber

Appendix G - 5 Performance of RBC with Angular Velocity 3.0 RPM and Rate of Aeration 5 SCFM

Appendix G - 6 Performance of RBC with Angular Velocity 4.0 RPM and Rate of Aeration 5 SCFM

Appendix G - 7 Performance of RBC with Angular Velocity 4.0 RPM and Rate of Aeration 8 SCFM

Appendix G - 8 Performance of RBC with Angular Velocity 2.5 RPM and Rate of Aeration 10 SCFM

Appendix G - 9 Performance of RBC with Angular Velocity 3.0 RPM and Rate of Aeration 10 SCFM

Appendix G - 10 Performance of RBC with Angular Velocity 4.0 RPM and Rate of Aeration 10 SCFM

Appendix G - 11 Performance of RBC with Angular Velocity 5.0 RPM and Rate of Aeration 10 SCFM

Appendix G - 12 Performance of RBC with Angular Velocity 4.0 RPM and Rate of Aeration 12 SCFM

Appendix G - 13 Performance of RBC with Angular Velocity 2.5 RPM and Rate of Aeration 15 SCFM

Appendix G - 14 Performance of RBC with Angular Velocity 3.0 RPM and Rate of Aeration 15 SCFM

Appendix G - 15 Performance of RBC with Angular Velocity 4.0 RPM and Rate of Aeration 15 SCFM

Appendix G - 16 Performance of RBC with Angular Velocity 5.0 RPM and Rate of Aeration 15 SCFM

Appendix H. Statistical Analysis – Optimum Parameters for Minimum Unit Energy Consumption

A. Unit Energy Consumption (KWH/kg CODr)

x = Angular Velocity, RPM; y = Rate of Aeration, SCFM; z = Unit Energy Consumption, KWH/kg CODr

Rank 1 Eqn 1022 :

$$z = \frac{a + bx + cy + dy^2 + ey^3}{1 + fx + gx^2 + hy + iy^2 + jy^3}$$

R^2 (Coefficient of Determination)	DF Adj R^2	Fit Std. Error	F-value
0.9948	0.9844	0.0673	127.95

Parameter	Value	Std Error	t-value	95% Confidence Limits	
a	0.2089	0.1828	1.1431	-0.2402	0.6580
b	-0.0038	0.0503	-0.0751	-0.1275	0.1199
c	-0.2624	0.3750	-0.6998	-1.1839	0.6590
d	0.0342	0.0687	0.4987	-0.1345	0.2030
e	-0.0014	0.0031	-0.4700	-0.0090	0.0061
f	-0.2576	0.2409	-1.0692	-0.8496	0.3344
g	0.0318	0.0290	1.0964	-0.0395	0.1032
h	-0.2899	0.2673	-1.0829	-0.9463	0.3673
i	0.0326	0.0449	0.7258	-0.0777	0.1429
j	-0.0012	0.0019	-0.6217	-0.0058	0.0035

X at Fn Zmin	Y at Fn Zmin	Fn Zmin
5	2.300	-6.98397e+10
X at Fn Zmax	Y at Fn Zmax	Fn Zmax
2.5	15	2.0595

Method: Gaussian Elimination

R^2 (Coefficient of Determination)	DF Adj R^2	Fit Std. Error
0.9948	0.9844	0.0673

Source	Sum of Squares	DF	Mean Square	F
Regr	5.2131	9	0.5792	127.95
Error	0.0272	6	0.0045	
Total	5.2402	15		

	Min	Max	Range	Mean	Std. σ
x	2.5	5	2.5	3.65625	0.8892
y	0	15	15	8.125	5.7835
z	0.3381	2.0341	1.6960	1.3068	0.5912

B. Unit Energy Consumption (KWH/kg BODr)

x = Angular Velocity, RPM; y = Rate of Aeration, SCFM; z = Unit Energy Consumption, KWH/kg BODr

Rank 1 Eqn 1022 :
$$z = \frac{a + bx + cy + dy^2 + ey^3}{1 + fx + gx^2 + hy + iy^2 + jy^3}$$

R ² (Coefficient of Determination)	DF Adj R ²	Fit Std. Error	F-value
0.9948	0.9844	0.1035	127.95

Parameter	Value	Std Error	t-value	95% Confidence Limits	
a	0.3214	0.2812	1.1431	-0.3695	1.0123
b	-0.0058	0.07742	-0.0751	-0.1961	0.1845
c	-0.4037	0.5770	-0.6998	-1.8214	1.0139
d	0.0527	0.1056	0.4987	-0.2069	0.3122
e	-0.0022	0.0047	-0.47007	-0.0138	0.0093
f	-0.2576	0.2409	-1.06927	-0.8496	0.3344
g	0.0318	0.0290	1.09637	-0.0395	0.1032
h	-0.2895	0.2673	-1.0829	-0.9463	0.3673
i	0.0326	0.0449	0.7258	-0.0777	0.1429
j	-0.0012	0.0019	-0.6217	-0.0058	0.0035

X at Fn Zmin	Y at Fn Zmin	Fn Zmin
5	2.3001	-1.07807e+11
X at Fn Zmax	Y at Fn Zmax	Fn Zmax
2.5	15	3.1685

Method: Gaussian Elimination

R ² (Coefficient of Determination)	DF Adj R ²	Fit Std. Error
0.9948	0.9844	0.1035

Source	Sum of Squares	DF	Mean Square	F
Regr	12.3386	9	1.3710	127.95
Error	0.0643	6	0.0107	
Total	12.4029	15		

	Min	Max	Range	Mean	Std. σ
x	2.5	5	2.5	3.65625	0.8891709622
y	0	15	15	8.125	5.7835974964
z	0.5202	3.1294	2.6093	2.0105	0.9093

C. Unit Energy Consumption (KWH/kg UODr) - Using Eqn. 1022

x = Angular Velocity, RPM; y = Rate of Aeration, SCFM; z = Unit Energy Consumption, KWH/kg UODr

Rank 1 Eqn 1022 :
$$z = \frac{a + bx + cy + dy^2 + ey^3}{1 + fx + gx^2 + hy + iy^2 + jy^3}$$

R^2 (Coefficient of Determination)	DF Adj R^2	Fit Std. Error	F-value
0.9935	0.9805	0.0690	101.98

Parameter	Value	Std Error	t-value	95% Confidence Limits	
a	0.2042	0.1859	1.0983	-0.2527	0.6611
b	-0.0057	0.0512	-0.1111	-0.1315	0.1201
c	-0.2347	0.3789	-0.6194	-1.1658	0.6964
d	0.0292	0.0682	0.4285	-0.1383	0.1967
e	-0.0011	0.0030	-0.3857	-0.0084	0.0061
f	-0.2463	0.2565	-0.9601	-0.8765	0.3840
g	0.0300	0.0308	0.9748	-0.0456	0.1056
h	-0.2923	0.3024	-0.9669	-1.0353	0.4506
i	0.0323	0.0505	0.6393	-0.0918	0.1564
j	-0.0011	0.0021	-0.5302	-0.0063	0.0041

X at Fn Zmin	Y at Fn Zmin	Fn Zmin
5	2.3191	-2.13649e+10
X at Fn Zmax	Y at Fn Zmax	Fn Zmax
2.5	15	1.9299

Method: Gaussian Elimination

R^2 (Coefficient of Determination)	DF Adj R^2	Fit Std. Error
0.9935	0.9805	0.0690

Source	Sum of Squares	DF	Mean Square	F
Regr	4.3750	9	0.4861	101.98
Error	0.0286	6	0.0048	
Total	4.4036	15		

	Min	Max	Range	Mean	Std. σ
x	2.5	5	2.5	3.65625	0.8892
y	0	15	15	8.125	5.7836
z	0.3085	1.8884	1.5800	1.1969	0.5418