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Overcoming Technological Challenges for the Commercialization of the Circulating Fluidized Bed Bioreactor for Municipal Wastewater Treatment

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Abstract

The fluidized bed bioreactor as an attached growth wastewater treatment process has demonstrated advantages over suspended growth processes for municipal wastewater treatment applications. However, previous studies have also demonstrated potentially serious disadvantages in terms of energy consumption and maximum reactor size of high flow applications.

In this work, a cost analysis using the CapdetWorks, supplemented by calibrated model data taken from GPS-X was performed to determine the cost effectiveness of the circulating fluidized bed bioreactor (CFBBR). This study demonstrated that the CFBBR is most cost competitive at low flow below 5 MGD. A 10%-20% reduction in net present values on a 30-year basis was estimated for the circulating fluidized bed bioreactor at flows of 5 MGD and lower, and similar costs (<10%) at 10 MGD and above. The study also showed that liquid pumping, aeration, and chemical consumption for phosphorus removal were major contributor to the OpEx and net present value.

The cost analysis identified small scale wastewater markets as the best for the CFBBR. One of the largest of such markets is rural China, with over half a billion people living in rural villages in China (and only the capacity to treat 3% of their sewage generated). A study on a pilot-scale twin fluidized bed bioreactor system was conducted in Guangzhou, China, treating a septic tank effluent from a residential building. The TFBBR, demonstrated COD and nitrogen removal rates of 92% and 82%, respectively. It further demonstrated a low biosolids production, corresponding to a cost for biosolids management roughly 50% that of a typical suspended growth treatment process. A cost comparison of estimates for COD addition (to facilitate denitrification) and biosolids treatment for the TFBBR and a conventional attached growth process showed that the TFBBR would be less expensive than conventional processes

To explore the energy and cost saving potential of inverse fluidization for the circulating fluidized bed bioreactor, several expanded mineral materials were tested as carriers for inverse three-phase fluidized bed bioreactors. After overcoming operational challenges, expanded clay as a carrier demonstrated good COD and ammonia removal efficiencies (93% and 98%, respectively) at loadings of 2.2 $kgCOD/m^3/d$ and 0.2 $kgN/m^3/d$, similar to previous studies on the inverse fluidized bed bioreactors. However, the observed high suspended biomass concentration indicated that clay could not operate strictly as an attached-growth process, but instead more as a hybrid process of attached and suspended growth.

Accurate methods for estimating liquid velocity would be a key tool for the design of fluidized bed bioreactors, enabling precise delineation of energy demands. Different methods for estimating bed voidage by particles properties and liquid velocity were explored. For low density and low Archimedes number particles, the Khan and Richardson correlation for estimating the n-index of the Richardson-Zaki equation was shown accurate within an average error of ± 3%. Furthermore, using Karamanev's correlation for the drag coefficient coupled with Newton's equation for the terminal velocity of free settling particles was accurate within ± 10% error.

Keywords

Fluidized bed bioreactor, Biological Wastewater Treatment, Biological Nutrient Removal, CapdetWorks Cost-Benefit Analysis, Inverse Fluidization, Bed Voidage

Lay Audience Summary

The circulating fluidized bed bioreactor is a process in which a biofilm of bacteria and other microbes is grown on particles to form bioparticles. The bioparticles are then "fluidized" or mixed by air flow or liquid circulation in the system.

One promising application of this technology is biological wastewater treatment. Typical wastewater treatment processes use suspended, or free-swimming, bacteria to consume organic and nutrient pollutants in wastewater. However, these processes require a great deal of space, energy, and money to build and operate. They also produce a lot of excess waste sludge that must be treated before final disposal.

Previous studies on the circulating fluidized bioreactor have shown that it is capable of treating wastewater at significantly higher rates than conventional processes using smaller reactor sizes; ideal for reducing the land use footprint. Furthermore, it also produces far less waste sludge compared to conventional processes. However, the circulating fluidized bed bioreactor does have a high energy footprint due to its air and water circulation needs. Whether this energy footprint is enough to offset its potential cost saving advantages is the purpose of this research work.

Costing estimates of the CFBBR and competing technologies were generated using a software called CapdetWorks. These estimates provided a detailed look at the cost of individual components of the treatment plants using the different technologies and allowed for the identification of major cost contributors. This provided two major findings, one) where the CFBBR currently stands in terms of cost competitiveness and 2) which areas of the CFBBR process should receive more attention by researchers in order to further improve its cost competitiveness.

One the major cost contributors was the fluidization energy, specifically liquid pumping. To explore reducing this cost, two studies on inverse fluidization (using floating particles instead of settling ones) were done. One study explored methods for predicting liquid pumping requirements based on the particle properties, evaluating several literature sources to determine which was the most accurate. The other study evaluated expanded minerals particles for use in inverse fluidized bed bioreactors and determine their treatment capability for municipal wastewater.

Co-Authorship Statement

This PhD thesis contains materials has been published, is under review, or in preparation for submission to peer reviewed journals. Specific details listed below.

Chapter 2: Fluidized-Bed Bioreactor Applications for Biological Wastewater Treatment: A Review of Research and Developments

Michael J. Nelson, George Nakhla, Jesse Zhu

Published in *Engineering*

The primary of author of this chapter was Michael J. Nelson under the supervision of Dr. Nakhla and Dr. Zhu. The literature review was conducted by Michael J. Nelson with guidance from Dr. Nakhla and Dr. Zhu, who also provided feedback and revisions to the manuscript.

Chapter 3: Decentralized wastewater treatment in an urban setting: a pilot study of the circulating fluidized bed bioreactor treating septic tank effluent

Anqi Liu, Michael J. Nelson, Xiaobo Wang, Haibin Li, Xiaoqin He, Zengli Zhao, Huiqiong Zhong, George Nakhla & Jesse Zhu

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The primary author of this chapter was Anqi Liu. The experimental plan and execution of experiments and data collection was conducted by Anqi Liu. Data analysis and drafting of the manuscript was done by Michael J. Nelson under the supervision of Dr. Zhu and Dr. Nakhla. Feedback and revisions on the manuscript was provided primarily by Anqi Liu, Dr. Zhu and Dr. Nakhla as well as from other co-authors.

Chapter 4: The Circulating Fluidized Bed Bioreactor as a Biological Nutrient Removal Process for Municipal Wastewater Treatment: Process Modelling and Costing Analysis

Michael J. Nelson, George Nakhla, Jesse Zhu

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The primary author of this chapter was Michael J. Nelson under the supervision of Dr. Nakhla and Dr. Zhu. The modelling and simulation plan was devised by Michael J. Nelson and Dr. Nakhla. Modelling, simulation, data analysis, and drafting of the manuscript was conducted by Michael J. Nelson. Feedback and revisions on the manuscript was provided by Dr. Nakhla and Dr. Zhu.

Chapter 5: Expanded mineral materials as carrier media in an inverse three-phase fluidized bed bioreactor for wastewater treatment

Michael J. Nelson, George Nakhla, Jesse Zhu

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Chapter 6: Bed Voidage Predictions for Inverse Liquid-Solid Fluidized Beds

Michael J. Nelson, Saleh Srabet, Tian Nan, Dominic Pjontek, Jesse Zhu

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List of Abbreviations and Symbols

Abbreviations

CFBBR – circulating fluidized bed bioreactor

TFBBR – twin fluidized bed bioreactor

LSCFB – liquid-solid circulating fluidized bed

FBBR – fluidized bed bioreactor

AnFBR – anaerobic fluidized bed bioreactor

BNR – biological nutrient removal

MWW – municipal wastewater

COD – chemical oxygen demand

TCOD – total chemical oxygen demand

SCOD – soluble chemical oxygen demand

BOD – biochemical oxygen demand

TN – total nitrogen

TP – total phosphorus

BPR – biological phosphorus removal

PAO – polyphosphate accumulating organism

TSS – total suspended solids

VSS – volatile suspended solids

PS – primary sludge

TWAS – thickened waste activated sludge

DO – dissolved oxygen

OLR – organic loading rate

HRT – hydraulic retention time

SRT – solids retention time

EBCT – empty bed contact time

Symbols

Ar	Archimedes number, $gl^3\rho_l(\rho_s - \rho_l)/\mu^2$ [-]
Re	Reynolds number at liquid velocity, $\rho_l d_p U_l/\mu$ [-]
Re_t	Reynolds number at terminal particle velocity, $\rho_l d_p U_t/\mu$ [-]
C_D	Drag coefficient of a particle in fluidized bed [-]
D	Column reactor diameter [m]
d_p	Particle diameter [m]
F_B	Buoyancy force [N]
F_D	Drag force [N]
F_G	Gravitational force [N]
g	Gravitational acceleration, [m/s^2]
n	Exponent in Richardson and Zaki correlation [-]
m	Particle mass [kg]
U_l	Superficial liquid velocity [m/s]
U_t	Terminal particle velocity [m/s]
h_m	Different liquid height between two manometers [m]
h_1	Liquid height in manometer [m]
Δh	Distance between the pressure ports [m]
ID	Internal column diameter [m]
AARE	Average Absolute Relative Error

Greek Letters

ρ_l	Liquid density [kg/m^3]
ρ_p	Solid density [kg/m^3]
μ_l	Liquid viscosity [$N.s/m^2$]
ε	Friction of bed volume by liquid [-]
ε_s	Friction of bed volume by solid particle (solids holdup) [-]

Chapter 1

1. Introduction

1.1 Rationale

With the world's population growing, the production of wastewater and subsequent demand for wastewater treatment is ever increasing, being crucial for the protection of aquatic ecosystems and drinking water sources(Metcalf and Eddy, 2003). Expanding urbanization is putting further pressure on urban municipal treatment plants, as they must not only meet increasingly stringent discharge standards with limited space and financial resources but also increase processing capacity. Thus, process intensification is becoming increasingly important for treatment plant upgrades, to minimize space requirements. Smaller cities, towns and rural municipalities also face increasingly stringent discharge standards, especially nutrient discharge standards for nitrogen and phosphorus. Such plants will benefit from less expensive biological nutrient removal processes that can meet these treatment requirements. In addition to process size, nutrient removal performance, and effluent quality, other factors that are important for wastewater treatment process selection are energy consumption, chemical usage, and the management and disposal of waste biosolids. All of the aforementioned factors contribute to a growing demand for a more compact, cost-effective biological treatment processes.

The circulating fluidized bed bioreactor (CFBBR) has been demonstrated as a biological nutrient removal process capable of handling high organic and nutrient loads compared to conventional suspended and attached growth biological process like activated sludge, integrated fixed-film activated sludge (IFAS) and moving bed bioreactors (MBBR)(Andalib et al., 2010; Chowdhury et al., 2009a, 2009b). It has also been demonstrated to be adaptable to several different biological treatment processes, namely, nitrification, denitrification, anaerobic digestion of biosolids, and anaerobic treatment of high-strength organic wastewater(Andalib et al., 2012; Islam et al., 2012; Mustafa et al., 2014) . For all applications, the CFBBR handles considerably higher loads than its competitors and produces less biosolids during treatment. Conversely, studies on the CFBBR have also shown that it has a higher energy demand due to its fluidization requirements and can also require more aeration for aerobic processes due to its low biosolids generation(Nelson et al., 2017). Limitations on the maximum size of fluidized bed columns will also inflate the capital cost due to increased modularity of the CFBBR at high flows. However, all of these potential advantages and limitations remain unexplored due to a lack of large-scale system data and limited methods for estimating large-scale costs for the CFBBR (or any process in early development).

Process modelling and simulation can be used to predict the performance of scale-up systems when calibrated with lab and pilot-scale data(Abbasi et al., 2021; Ghasemi et al., 2020). This will provide valuable insight into how new processes will compare with their competitors in terms of meeting treatment standards and demand. However, process modelling and costing softwares (BioWin, GPS-X, CapdetWorks) generally do not have process models for newer, emerging technologies(Abbasi et al., 2021; McGhee et al., 1983; Pineau et al., 1985; Wright et al., 1988). This presents a challenge of developing a calibrated CFBBR model that can be used for treatment simulation and cost estimation. The estimation of technology cost is almost equally important to its scientific/technical aspects development, as with detailed, accurate costing estimates of emerging technologies, not only would their effects on the overall treatment costs be delineated, but also specific aspects of the process that contribute more to the cost can be identified for further study and optimization.

1.2 Objectives

The CFBBR has been demonstrated as a potential high-rate alternative for conventional biological treatment processes, but no scale-up cost analysis has been done thus far to determine its economic viability. With accurate cost predictions, areas for optimization can be identified to improve the CFBBR's cost competitiveness. The specific objectives of this thesis are:

1) Evaluate current treatment capabilities of the CFBBR in terms of unit size, process loadings, individual applications, and effluent quality.
2) Develop a calibrated process model for the CFBBR in GPS-X for comparative performance evaluation and cost analysis.
3) Determine flow ranges and applications for which the CFBBR is cost-competitive in order to identify the best markets for commercialization efforts, and the areas for further technical/scientific development.
4) Identify hinderances of the CFBBR's cost competitiveness both in terms of treatment performance and economic competitiveness and explore pathways to improve the cost effectiveness of the CFBBR.

1.3 Thesis Organization

Chapter 1 provides a summary of the thesis and rationale for studying the CFBBR potential applications and cost effectiveness. This chapter also provides the specific research objectives for this body of work.

Chapter 2 is a review article entitled "*Fluidized-Bed Bioreactor Applications for Biological Wastewater Treatment: A Review of Research and Developments.*" This article provides a detailed review of the research-to-date (circa 2017) on the CFBBR. It covers different applications for which it has been tested, including mainstream wastewater treatment and biological nutrient removal, anaerobic digestion of biosolids, and anerobic treatment of high-strength organic wastewaters. The review also covers process modelling studies conducted on the CFBBR using the softwares AQUIFAS and BioWin. The major conclusions of this chapter are that the CFBBR has been clearly demonstrated to achieve superior treatment to its competitors in terms of process loadings and effluent quality and is also a very adaptable platform for process development, for a range of aerobic, anoxic and anaerobic biological treatment processes.

Chapter 3 is a research article entitled "*Decentralized wastewater treatment in an urban setting: a pilot study of the circulating fluidized bed bioreactor treating septic tank effluent.*" The objective of this study was to evaluate the performance of a pilot-scale CFBBR treating the effluent of a residential septic tank with a low COD:N ratio, necessitating carbon supplementation for nitrogen removal via denitrification. Three stages of treatment were tested: one stage without supplemental carbon and two with different amounts of supplemental carbon. The supplemental carbon requirements and biosolids generation were compared to a conventional activated sludge-based process, as biosolids generation is a major cost contributor and supplemental carbon increases the amount of biosolids. Therefore, it is important to optimize the supplemental carbon to achieve complete nutrient removal while also minimizing the amount of biosolids produced.

Chapter 4 is a research article entitled "*The Circulating Fluidized Bed Bioreactor as a Biological Nutrient Removal Process for Municipal Wastewater Treatment: Process Modelling and Costing*

Analysis." The main purpose of this work was to develop a method to accurately predict the scaled-up costs of the CFBBR, assess its economic competitiveness and identify areas for optimization in terms of both treatment and economic performance. The specific objective of this work was to perform a cost benefit analysis of the CFBBR against other competing technologies in the context of a "whole-plant" cost simulation. To this end, a calibrated CFBBR process model developed in GPS-X was used in conjunction with the costing software CapdetWorks (both of Hydromantis, Inc) to generate detailed capital and O&M cost estimates. Both software packages were also used to model and generate costing estimates for competing technologies treating the same simulated wastewaters. The costing estimates were validated with real life data from the US EPA relevant database. The cost estimates for various technologies were compared in terms of total CapEx (capital), OpEx (O&M), and net present value. Cost analysis of individual components of the plants helped delineate the major drivers for further CFBBR research.

Chapter 5 is a research chapter entitled "*Expanded mineral materials as carrier media in an inverse three-phase fluidized bed bioreactor for wastewater treatment*." Expanded light weight mineral-type carrier were evaluated as carrier media for inverse fluidized bed bioreactor applications. This research showed numerous operational challenges with most carriers tested due to biofilm and water absorption affecting the fluidization characteristics of the particles. Only expanded clay was able to maintain fluidization and achieve nitrification. However, the carrier showed a high degree of biofilm detachment, meaning that a strictly attached growth process could not be maintained. Instead, the system operated as a hybrid of attached and suspended growth. Nonetheless, the system was able to achieve COD and ammonia removal rates comparable to previously studied inverse fluidized bed applications. The presence of high suspended growth, though disadvantageous for an attached process, presents an opportunity to the explore the use of expanded as a carrier media in a hybrid process designed for enhanced biological nutrient removal, as the suspended biomass can be circulated between aerobic and anoxic columns much more easily than circulating the media.

Chapter 6 is a research article entitled "*Bed Voidage Predictions for Inverse Liquid-Solid Fluidized Beds*." As inverse fluidization offers a potential energy and cost saving means for the CFBBR, the purpose of this work was to explore the hydrodynamics of inverse fluidization. The objective of this work was to evaluate multiples methods for estimating bed expansion of inverse fluidized beds based on particle properties and liquid velocity. In this study, the solids holdup of four sizes of Styrofoam particles was measured and several correlations from literature for predicting the n-index of the Richardson-Zaki equations and predicting terminal velocity were evaluated using the data obtained.

Chapter 7 provides the major conclusions and scientific contributions of the thesis and offers perspectives on the future research to be conducted on the CFBBR.

1.4 Thesis Format

This thesis is written is the integrated article format in accordance with the specifications put forth by the School of Graduate and Post-Doctoral Studies at the University of Western Ontario. Chapter 2 of this thesis has been published in the journal *Engineering*. Chapter 3 has been published in the journal *Environmental Technology*. Chapter 4 has been accepted for publication in the *Journal of Environmental Management*. Chapter 5 is not being prepared for publication at this time. Chapter 6 is prepared to be submitted to the journal *Particuology*.

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Chapter 2
Fluidized Bed Bioreactor Applications for Biological Wastewater Treatment: A Review of Research and Developments

Abstract

Wastewater treatment is a process vital to protecting both the environment and human health. Currently, the most cost-effective way of treating wastewater is with biological treatment processes, such as the activated sludge process, albeit their long operating times. However, the population increase will create a demand for more efficient means of wastewater treatment. Fluidization has demonstrated its ability to increase the efficiency of many processes in chemical and biochemical engineering, but it has not been widely used in large-scale wastewater treatment. At Western University, the Circulating Fluidized Bed Bioreactor (CFBBR) was developed for treating wastewater. In this process, carrier particles develop a biofilm composed of bacteria and other microbes. The excellent mixing and mass transfer characteristics inherent to fluidization made this process very effective at treating both municipal and industrial wastewater. Studies of lab and pilot scale systems showed that the CFBBR could remove over 90% of the influent organic matter, 80% of the nitrogen and produce less than one-third of the biological sludges of the activated sludge process. Due to its high efficiency, it is also able to treat wastewaters with high organic solid concentrations, which are more difficult to treat with conventional methods because they require longer residence times, thus reducing the system size and footprint. It is also much better at handling and recovering from dynamic loadings (varying influent volume and concentrations) than current systems. Overall, the CFBBR has proven to be a very effective means of treating wastewater, capable of treating larger volumes of wastewater using a smaller reactor volume and a shorter residence time, with its compact design opening potential for more geographically localised and isolated wastewater treatment systems.

2.1 Introduction

2.1.1 Biological Wastewater Treatment

Wastewater treatment is an important process for protecting both the environment and human health. Pollutants and bacteria in wastewater can cause severe damage to water resources, which can in turn be damaging to humans and other animals that come in contact with it. For centuries, humans could simply release their waste into environment with little or no effect, as nature was able to take up the pollutants. But as the population grew, nature's capacity for taking up the pollutants was exceeded and the wastewater needed treatment before release or risk damaging nature and humans. Now, wastewater is collected from buildings (residential, industrial, business, medical, etc.) and enters sewer pipe systems. It then flows through the pipes and pumping stations (to keep the flow moving) until it reaches a treatment plant. There are now many established processes that are capable of treating wastewater effectively. However, as the human population continues to increase, more wastewater will be produced creating a larger demand for treatment (Metcalf and Eddy, Inc, 2003).

To meet the ever-increasing demand of treating wastewater, new plants will need to be constructed and existing plants will require upgrades and expansions. These new plants and expansions will take up more space in the population centres and as cities expand, less space will be available for the treatment plants. To combat this, more efficient treatment processes capable of treating larger volumes of wastewater in less time than the conventional methods will be needed. One technology that has been proven to have a high efficiency for treating wastewater is the circulating fluidized bed bioreactor (CFBBR) (Cui, Nakhla, Zhu, & Patel, 2004) developed at Western University, Canada. This review will cover the research done at Western University on the lab and pilot scale systems treating municipal wastewater and various industrial wastewaters.

The primary pollutants that must be removed from wastewater are carbon, nitrogen, and phosphorus. This includes organic compounds, ammonia, phosphates, and many others. Particulate and colloidal solids must also be removed. Finally, harmful pathogens need to be stabilized and/or destroyed (Metcalf and Eddy, Inc, 2003).

The conventional layout of a wastewater treatment plant (WWTP) starts with primary treatment, which removes large solids through physical separation process like screening and gravity settling, followed by secondary treatment, where most of the biological treatment occurs. Finally, the wastewater moves to tertiary treatment, where it is chemically polished and disinfected (if necessary). **Figure 2-1** shows the basic layout of a WWTP that uses an activated sludge system.

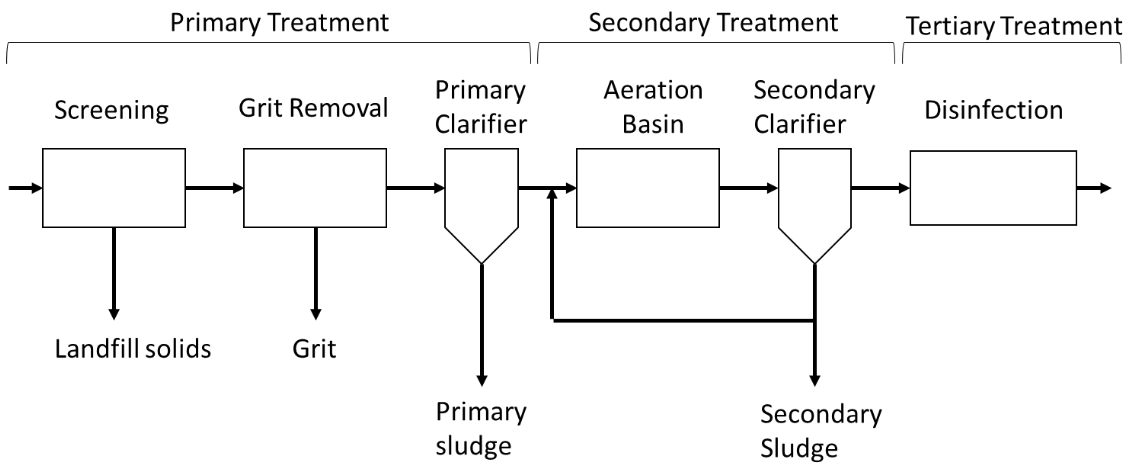

Figure 2-1 Layout of a conventional WWTP

Biological treatment processes are employed in the secondary treatment section. Biological treatment is carried out by microbial growth contained within bioreactors that consumes the pollutants through its metabolic processes. It generally comes in two forms: suspended and attached growth. Suspended growth has the microbial colonies (flocs) free swimming/floating in the mixed liquor. This mixing is induced either mechanically by impellers or by air flow from the bottom. The most well known of the suspended growth process is the activated sludge process, seen in **Figure 2-1**. Attached growth, also called fixed film, characterized by a biofilm composed of bacteria, particulates, extracellular polymers and gels growing on a support media, is shown in **Figure 2-2**. Typical carrier media used for attached growth is rock or plastic. Ideal carriers are porous and have rough surfaces because they allow more effective attachment compared to smooth, non-porous surfaces (Metcalf and Eddy, Inc, 2003).

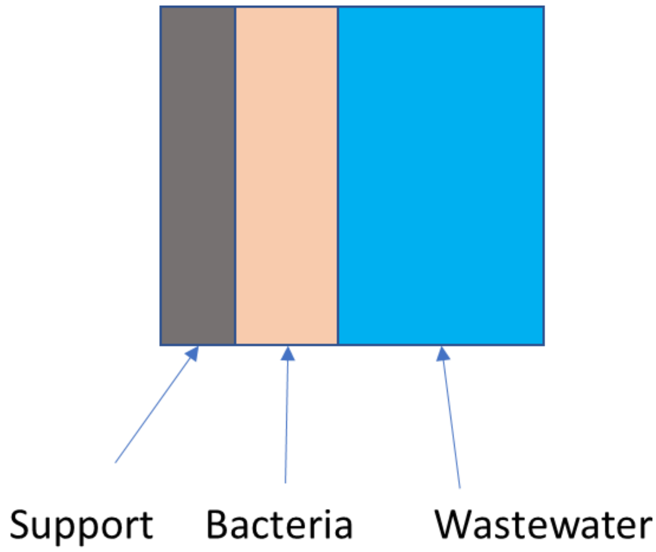

Figure 2-2 Attached growth

The four main processes carried out in general wastewater treatment are aerobic organic oxidation, nitrification, denitrification, and biological phosphorus removal. It is through these processes that most of the carbon, nitrogen, and phosphorus are removed. These are carried out by different types of bacteria and require different environmental conditions and substrates (Metcalf and Eddy, Inc, 2003). Two classes of bacteria to be considered are based on the type of carbon they consume for cell growth: heterotrophic, which consume organic carbon, and autotrophic, which consume inorganic carbon. The three environmental conditions to consider are aerobic (presence of oxygen), anoxic (presence of nitrates, low to no oxygen) and anaerobic (no oxygen or nitrates) (Metcalf and Eddy, Inc, 2003).

Aerobic Organic Oxidation

Heterotrophic bacteria oxidise organic material to gain energy and use it for biomass synthesis. The basic reaction is as follows:

$$C_{10}H_{19}O_3N \text{ (organic material)} + 12.5O_2 \rightarrow 10CO_2 + 8H_2O + NH_3 \quad (2\text{-}1)$$

As seen in reaction (2-1), the organic material ($C_{10}H_{19}O_3N$) is broken down to carbon dioxide, water and ammonia, using oxygen gas as the oxidizing agent (Metcalf and Eddy, Inc, 2003).

Nitrification

Autotrophic bacteria carry out a carbon fixation process using ammonia as the electron donor to convert inorganic carbon into organic carbon compounds. The reduction-oxidation (redox) reaction oxidizes the ammonia into nitrites and then nitrites to nitrates. This reaction is also used for gaining energy for other cellular functions. Due the lower growth yields and rates of these bacteria, most of the biodegradable organics (BOD) must be removed first. If not, the heterotrophic bacteria will dominate the growth and outcompete the nitrifying bacteria leading to washout of the nitrifiers.

$$NH_4^+ + HCO_3^- + O_2 + CO_2 \rightarrow NO_3^- + H_2O + C_5H_7O_2N \text{ (biomass)} \quad (2\text{-}2)$$

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \quad (32\text{-})$$

In reaction (2-2), nitrates are used as an electron donor to reduce inorganic carbon (HCO_3^- and CO_2) to organic carbon. Some of the ammonia is also incorporated into the new biomass. Reaction (2-3) shows the overall reaction of oxidizing ammonia into nitrates (Metcalf and Eddy, Inc, 2003).

Denitrification

Certain bacteria have a nitrate reductase enzyme in their electron transport chain that allow them to substitute nitrates for oxygen as the electron acceptor. Through this process, in a series of reactions, the nitrates are reduced to diatomic nitrogen which then bubbles out of the water due to its low solubility. It should be noted that this process can only occur in low oxygen and high nitrate concentrations (anoxic conditions), otherwise the nitrate reductase enzyme will be inhibited (Metcalf and Eddy, Inc, 2003).

$$10NO_3^- + C_{10}H_{19}O_3N \rightarrow 10CO_2 + 3H_2O + NH_3 + 5N_2 + 10OH^- \quad (2\text{-}4)$$

Reaction (2-4) is similar to reaction (2-1), in that organic material is oxidized into carbon dioxide, water, and ammonia. However, in this case, nitrates have replaced oxygen as the electron acceptor. When the nitrates are reduced, they become diatomic nitrogen (N_2) and leave the system as nitrogen gas (Metcalf and Eddy, Inc, 2003).

Enhanced Biological Phosphorus Removal

In addition to phosphorus removal by biomass synthesis, enhanced biological phosphorus removal (EBPR) is employed when large concentrations of phosphorus are present. EBPR is a two-stage process carried out by a group of bacteria called polyphosphorus accumulating organisms (PAO's). These microbes are capable of storing large amounts of phosphates in the form of polyphosphate granules. This is a method of energy storage to be used in place of adenosine triphosphate (ATP) when the aerobic metabolic pathways are not functional (absence of oxygen).

The first stage is an anaerobic process where the PAO's use their stored phosphate to take up and store organic material (acetate and short chain fatty acids), while simultaneously releasing phosphates into the water. The second stage is aerobic, where the PAO's use the stored fatty acids as an energy source for taking up the phosphates in the water and store them as polyphosphate granules. The microbes are then settled in the clarifier, recycled to the start of the process and any excess sludge is removed. In EBPR phosphates are ultimately removed in the waste sludge stored in the PAO's (Metcalf and Eddy, Inc, 2003).

2.1.2 Fluidized Bed Bioreactors

Fluidization is a process in which the upwards flow of a fluid suspends a bed of particles. Fluidization offers many advantages, the major ones being excellent mixing, increased mass transfer, large specific surface area, and uniform particle and temperature distributions. Fluidization was first used in the 1920's for coal gasification (Kunii & Levenspiel, 1991) and the second major application was fluidized catalytic cracking, developed in the 1940's (Jahnig, Campbell, & Martin, 1980). Both processes utilized gas-solid fluidization and it has since been developed and applied to many other processes. Later, liquid-solid, and three-phase fluidization were developed and have been proven to have great potential and application in biochemical processes (Zhu, Zhen, Karamanev, & Bassi, 2000). The basic outline and function of these two forms of fluidization will be covered in the next two sub-sections.

2.1.3 Fluidization

Liquid-solid (LS) fluidization works by an upwards moving liquid stream suspending and/or entraining a bed of solid particles. There are several fluidization regimes, but the two regimes that the wastewater processes to be discussed later operate within are the conventional and circulating regimes. **Figures 2-3 and 2-4** below show a basic depiction of conventional and circulating fluidization. In conventional fluidization, the liquid velocity is insufficient to entrain the particles and wash them out of the column (Kunii & Levenspiel, 1991). In circulating fluidization, a high liquid velocity is used to carry the particles to the top of the column and then returned to the bottom via a recycle line or column (Grace, 1990).

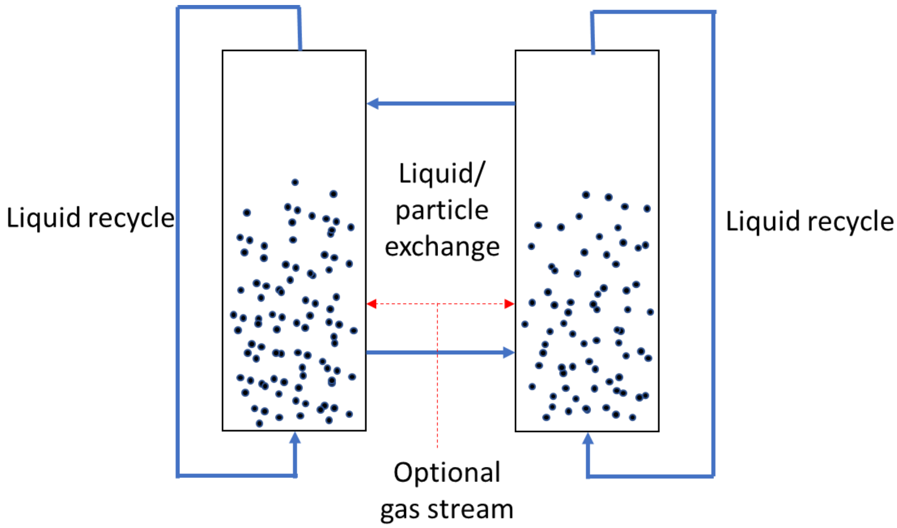

Figure 2-3 Conventional twin fluidized bed system (adapted from Zhu et al, 2000) *(Zhu, Zhen, Karamanev, & Bassi, 2000)*

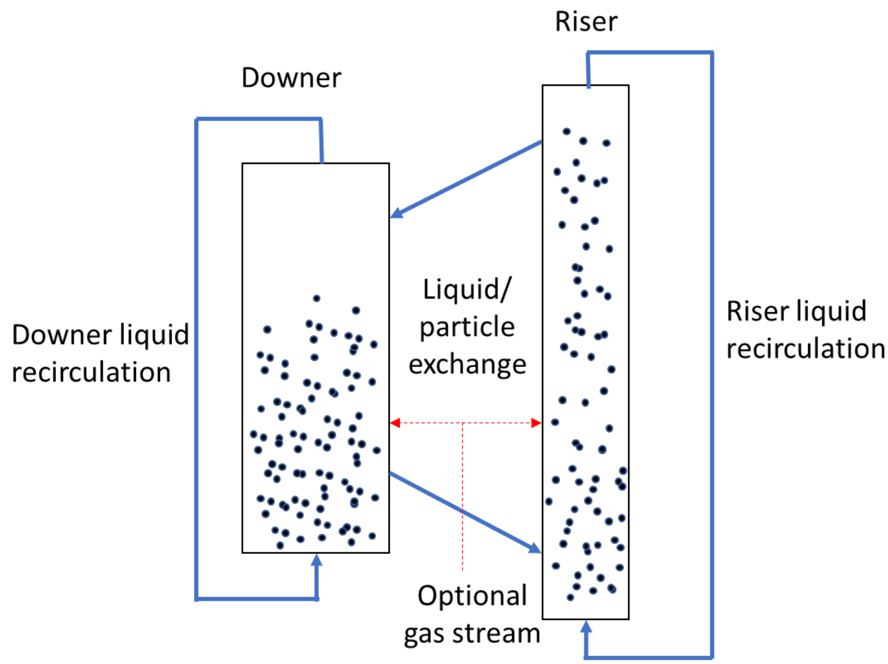

Figure 2-4 Layout of liquid-fluidized bed with particle circulation (adapted from Zhu et al, 2000) *(Zhu, Zhen, Karamanev, & Bassi, 2000)*

Three-phase (gas-liquid-solid (GLS)) has the same general layout as LS fluidization except with the addition of an air distributor as well as the liquid distributor. In GLS fluidization, both the liquid stream and gas bubbles fluidize the particles. Like LS fluidization, GLS fluidization can operate in both the conventional and circulating regimes (Zheng & Zhu, 1999). However, depending on the specific requirements of the process, only one of the columns may have a gas distributor and thus only one of the columns may operate with GLS fluidization, such as in the process to be discussed in this paper.

2.1.4 Principle of Fluidized Bed Bioreactor

The FBBR is an application of the liquid-solid fluidized bed. These bioreactors can be run in a single or double column system depending on the treatment process being carried out. The FBBR is an attached growth process. The microbes attach to the fluidized media and form a biofilm on the surface (**Figure 2-5**) (Metcalf and Eddy, Inc, 2003). Fluidization in the column is caused by the recirculating wastewater and/or the air stream, if the process includes aeration (Zhu, Zhen, Karamanev, & Bassi, 2000).

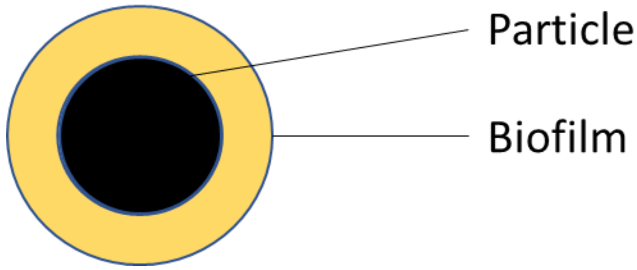

Figure 2-5 Particle-biofilm

Like all fluidization processes, the excellent mixing and increased mass transfer in the process enhance its function. The use of smaller particles compared to other attached growth systems like IFAS and MBBR coupled with excellent microbial attachment characteristics results in much thicker biofilms and hence, the surface area of the film exposed to the water is much higher than traditional attached growth processes. The increased contact between the wastewater substrates and the biofilm also allows it to breakdown larger compounds that are typically more difficult to treat. The FBBR has also been proven to be capable of handling larger loadings and operate at lower than typical hydraulic retention times.

2.2 Circulating Fluidized Bed Bioreactor (CFBBR)

The circulating fluidized bed bioreactor is the system developed at Western University. It is a twin column system. The twin column CFBBR is capable of maintaining two different environments within the system, which is advantageous for biological treatment (Zhu, Zhen, Karamanev, & Bassi, 2000). The CFBBR has an aerobic column (medium to high oxygen) and an anoxic column (low oxygen, high nitrates), which enables it to achieve nitrification and denitrification in the same process. It can also be run with particle exchange between the two columns enhancing phosphorus removal because of the transfer between aerobic and anaerobic environments (Patel, Zhu, & Nakhla, 2006). The CFBBR has been tested with lab and pilot scale reactors, treating municipal wastewater and leachate.

2.2.1 Scales of Research Studies

2.2.1.1 Lab Scale

CFBBR-1

The CFBBR was first tested with a lab scale reactor (CFBBR-1), consisting of a riser, downer, and liquid-solid separators at the top of each column. A schematic of the system is shown below in **Figure 2-6**. The configuration of CFFBR-1 is similar to that seen in **Figure 2-5** operating with particle circulation between the riser and downer. The system operated with the riser in the circulating fluidization regime and the downer in the conventional regime, with the liquid and particles being separated at the top in the LS separators. The particles at the top of the riser are transferred to the downer. Due to the tighter packing of the particles in the downer than in the riser, the biofilm-rich particles transferred to the downer from the riser will lose their biofilm due to shear and abrasion as the particles collide with each other. Accordingly, the loss of biomass increases the density of the particles, forcing them to move downwards through the conventional fluidized bed. The particles in the bottom of the downer are recirculated to the riser to begin the cycle again. The liquid at the top of the downer enters an LS separator, where most of the suspended solids (VSS and TSS) are separated for sludge wasting and the remaining nitrate-rich liquid is circulated back to the downer for fluidization and to the riser for fluidization and denitrification. In this apparatus, lava rock was used as the carrier media. The average particle diameter was 0.67 mm with a bulk and true density of 1720 kg/m^3 and 2560 kg/m^3, respectively. The lava rock had an approximate surface area of 9298 m^2/m^3 (Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009).

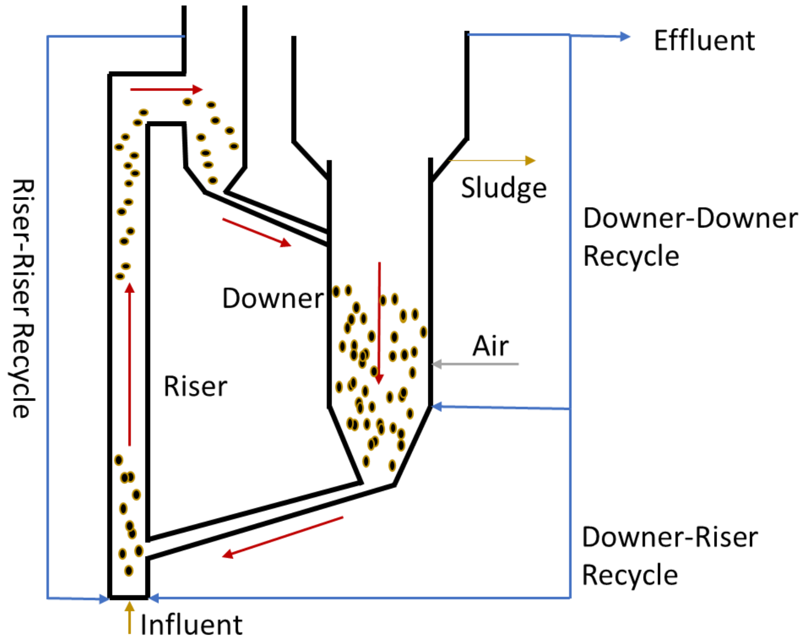

Figure 2-6 Diagram of CFBBR design and the directions of gas, liquid, and solid flow (Chowdhury et al, 2009) *(Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009)*

The system is designed so that the downer operates under aerobic conditions (three-phase fluidization) to achieve biological organic oxidation and nitrification of ammonia. The riser operates with anoxic conditions (two-phase fluidization) to achieve denitrification of nitrates. The system has also demonstrated potential for enhanced biological phosphorus removal but not to the same degree as a system designed specifically for EBPR, due to the CFBBR's lack of a true anaerobic zone which is required for EBPR (Metcalf and Eddy, Inc, 2003). A summary of its influent and effluent qualities is shown below in **Table 2-1**.

Table 2-1 Influent and Effluent Quality of CFBBR (Chowdhury et al, 2009) *(Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009)*

Parameter (mg/L)	Influent	Effluent
COD	273	26
SCOD	73	21
NH_4^+-N	19	0.7
NO_3^--N	0.5	6.5
TN	31.2	8.6
TP	3.8	0.8
TSS	144	4

VSS	118	3

CFBBR-2

The other system tested at lab scale is the Twin Fluidized Bed Bioreactor (CFBBR-2 or TFBBR). A diagram of this system is shown below in **Figure 2-7**. Like the CFBBR-1, the CFBBR-2 consists of two columns; one aerobic and one anoxic. However, these columns are the same height, and both operate with conventional fluidization, similar to the configuration shown in **Figure 2-3**. Since both columns operate in the conventional fluidization regime, there is no continuous particle exchange occurring. The TFBBR was designed after it was discovered that particle circulation does not play a significant factor in its treatment performance. Circulation of the particles is only necessary if enhanced phosphorus removal is required. Particle circulation between the riser and downer can be carried out with impellers at the top and bottom of the columns periodically transferring the particles, making particle circulation independent within the process. Particles at the bottom of the aerobic column would transferred to the anoxic column and those at the top of the anoxic column would be transferred to the aerobic column (Andalib, Nakhla, & Zhu, 2010). This particular system had two columns of identical shape and volume, however this is not necessary, as the columns' sizes can vary depending on the required HRT for each column.

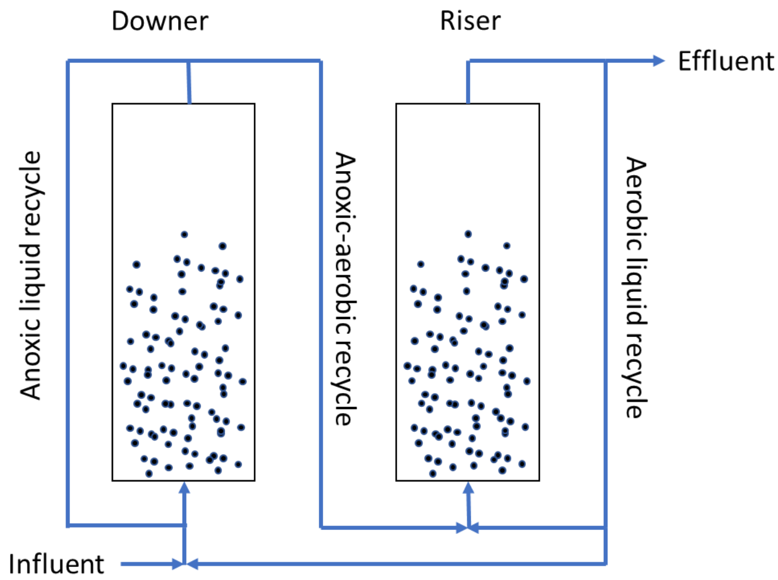

Figure 2-7 Diagram of the TFBBR System (Andalib et al, 2010)

As this system operated with convetional fluidization in both columns, the shear rate on the biofilm was lower the in the CFBBR-1. This led to a much lower detachment rate and longer SRT, culminating in a much lower observed biomass yield for the overall system. The observed solid yields ranged from 0.06-0.071 gVSS/gCOD (Andalib, Nakhla, & Zhu, 2010), which were significantly lower than the yields seen in CFBBR-1 (0.12-0.16 gVSS/gCOD) (Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009). In addition, the CFBBR-2 system was proven to have similar BNR performance and effluent quality to that of the CFBBR-1, which are sumarrized in **Tables 2-2** below.

Table 2-2 Influent and Effluent Quality of TFBBR (Andalib et al, 2010) *(Andalib, Nakhla, & Zhu, 2010)*

Parameter (mg/L)	Influent	Effluent
COD	262	20
SCOD	234	9.5
NH_4^+-N	26.1	0.5
NO_3^--N	0.7	3.9
TN	29.5	5.4
TP	4.4	3.8
TSS	27	16.3

VSS	19	12

The CFBBR-2 system was tested with synthetic wastwater at organic loading rates of 1.3, 1.7, and 2.3 kgCOD/m^3day. The effluent quality and BNR efficiency was similar for all of these loadings, with the effluent of 2.3 kgCOD/m^3day loading shown in **Table 2-2**. Above an OLR of 2.3 kgCOD/m^3day, the COD removal efficiency began to decrease due to increased shear on the particles and subsequent biomass detachment. However, in the lower OLR's the detachment rate was measured to be much lower than that of the CFBBR at both the lab and pilot scale, giving it a comparatively longer SRT (Andalib, Nakhla, & Zhu, 2010).

Effects of Particle Circulation – Enhanced Biological Phosphorus Removal

As observed in the circulating and twin-column platforms, the major difference is the occurrence of enhanced biological phosphorus removal (EBPR). EBPR occurs when poly-phosphorus accumuating organisms (PAOs) are transferred from anaerobic to aerobic environments. In the anaerobic stage, PAOs use stored phopshates in place of ATP for metalbolism and to accumulate energy stores in forms such as glycogen, while releasing phosphates into the bulk water. Then in the aerobic stage, PAOs take up phosphates into their biomass. Phosphorus is then removed via waste sludge from the secondary clarifiers(Metcalf and Eddy, 2003).

EBPR occurs in the circulating platofrm due to the circulation of biofilm-laden particles between the anoxic and aerobic columns. As attached biomass, via the particles, is exchanged between the anoxic and aerobic zones, EBPR occurs. However, there are multiple drawbacks to facilitating EBPR in this fashion(Metcalf and Eddy, 2003).

The PAOs must compete with denitrifiers for substrate in the anoxic column. This competition can be addressed by either extending the residence in the anoxic column, however, too long of an HRT in anoxic/anaerobic conditions will lead to fermentation occuring, which will reduce the effective of the PAOs. The other option is to have a dedicated anaerobic zone, as seen in the A2O and UCT suspended growth processes, which both have three sections; anaerobic, anoxic, and aerobic. Both employ a mixed-liquor recycle from the aerobic to anoxic zones in order to supply nitrates for denitrification independently of the return activated sludge. This greatly reduces the amount nitrates sent to the anaerobic zone, reducing PAO-denitrifier competition and increasing the effectiveness of the PAOs(Metcalf and Eddy, 2003). A dedicated anaerobic fluidized bed could be used for the CFBBR but this would contribute to higher capital costs for the additional column and particles, as well as higher operating costs from increased energy demands for fluidization and particle circulation between the three columns.

One other major disadvantage of EBPR in the fluidized bed platform is the significantly lower sludge yield. As the removal of phosphorus is directly connected to the removal of excess biomass, a lower sludge yield limits the amount of phosphorus that can be removed. This is why EBPR is typically only done deliberately in suspended growth processes, as suspended growth processes typically have higher biomass yields and shorter solids retention times than attached growth processes(N. Chowdhury et al., 2009a; N. Chowdhury, Nakhla, et al., 2010).

So, while the CFBBR can facilitate EBPR, it is limited in the amount of phosphorus it can remove due to competition between PAOs and denitrifiers limiting the PAOs effectiveness and the low sludge yield of the CFBBR limiting the total amount of phosphorus that can be removed with the waste biomass.

2.2.2 Pilot Scale

Following the success of the lab scale, a pilot scale system was established and tested at the Adelaide Wastewater Treatment Plant in London, Canada. As one of the City of London’s six wastewater treatment plants Adelaide treats an annual average of 27455 m^3/day (Environmental and Engineering Service Department, 2017). The system had the same general configuration, layout, and operation as the lab scale system. **Figure 2-8** below shows the design of pilot system used (Chowdhury, Nakhla, Zhu, & Islam, 2010).

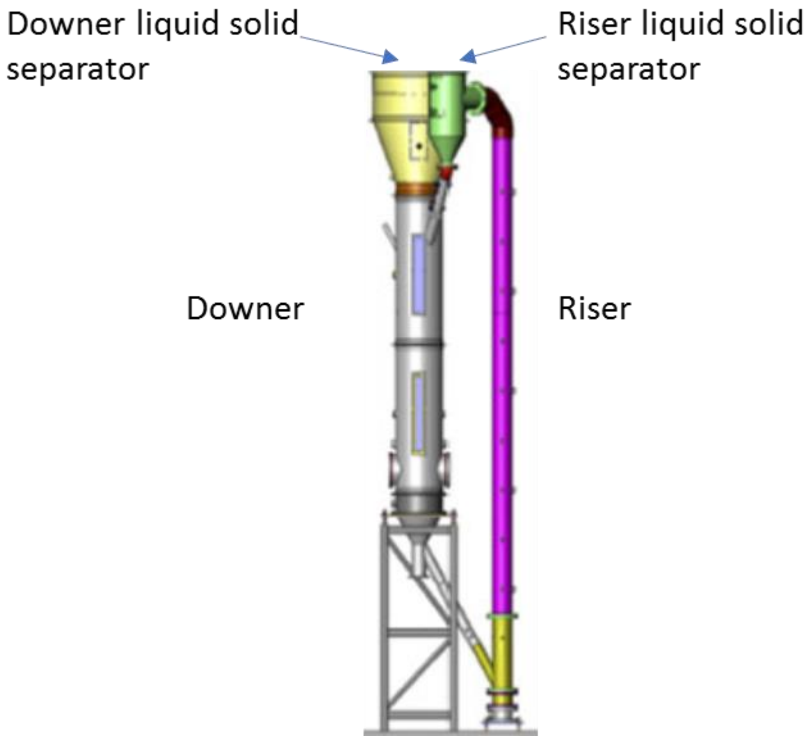

Figure 2-8 Configuration of Pilot Scale CFBBR (Chowdhury et al, 2010) *(Chowdhury, Nakhla, Zhu, & Islam, 2010)*

The carrier media for this system was also lava rock of similar particle diameter (average 0.67 mm) and density (1720 kg/m^3) to that used in lab scale system. The pilot scale CFBBR was designed to treat 5 m^3/d of primary influent and achieved removal efficiencies close to the lab scale system (Chowdhury, Nakhla, Zhu, & Islam, 2010). Another exceptional aspect of this system was demonstrated by the effluent quality, as the VSS and phosphorus concentrations were low enough to meet secondary effluent quality without the need for secondary clarification or chemical phosphorus removal. The ability to handle high solid feeds and produce low solid effluent could allow future plants to reduce the size and cost of clarifiers (Sutton & Mishra, 1990). **Table 2-3** shows the treatment data from all three phases of the study. The influent flow rates for phases 1-3 were 2880, 4320, and 5800 L/d on average, respectively. A full summary of the lab and pilot scale BNR efficiencies and those of alternative technologies and methods are shown is **Table 2-4** below.

Table 2-3 Influent and effluent data of pilot-CFBBR study (Chowdhury et al, 2010) *(Chowdhury, Nakhla, Zhu, & Islam, 2010)*

	Phase I (2880 L/d)	**Phase II (4320 L/d)**	**Phase III (5800 L/d)**

Parameter (mg/L)	Influent	Effluent	Influent	Effluent	Influent	Effluent
TCOD	332 $\pm$ 42	26 $\pm$ 3	349 $\pm$ 38	39 $\pm$ 8	496 $\pm$ 152	45 $\pm$ 7
SCOD	71 $\pm$ 14	13 $\pm$ 4	100 $\pm$ 16	15 $\pm$ 4	117 $\pm$ 23	23 $\pm$ 5
NH_4^+-N	22.1 $\pm$ 5.2	1.2 $\pm$ 0.5	24.6 $\pm$ 2.9	0.9 $\pm$ 0.3	25.8 $\pm$ 1.1	9.5 $\pm$ 0.9
NO_3^--N	0.9 $\pm$ 0.6	3.6 $\pm$ 1.2	0.4 $\pm$ 0.1	4.7 $\pm$ 1.3	0.4 $\pm$ 0.1	2.8 $\pm$ 0.6
TP	4.9 $\pm$ 1.0	1.0 $\pm$ 0.1	4.2 $\pm$ 0.8	1.2 $\pm$ 0.2	5.9 $\pm$ 0.6	1.2 $\pm$ 0.4
TSS	217 $\pm$ 27	11 $\pm$ 2	219 $\pm$ 26	22 $\pm$ 6	443 $\pm$ 174	27 $\pm$ 6
VSS	174 $\pm$ 28	9 $\pm$ 2	171 $\pm$ 23	16 $\pm$ 5	315 $\pm$ 106	21 $\pm$ 6

Table 2-4 Summary of BNR Performance

Name	Source	HRT (hrs)	EBCT (hrs)	SRT (days)	OLR (kg/m^{3}d)	COD (%)	N (%)	P (%)	Biomass Yields (mg_{VSS}/mg_{COD})
CFBBR-1	(Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009)	2.04	0.82	44-56	3.36	91	78	85	0.12-0.135
CFBBR-2/ TFBBR	(Andalib, Nakhla, & Zhu, 2010)	2.88	0.98	72-108	2.23	97	84	12	0.071
Pilot-CFBBR	(Chowdhury, Nakhla, Zhu, & Islam, 2010)	2.03	1.5	20-39	4.12	90	80	70	0.12-0.16
UASB	(La Motta, Silva, Bustillos, Padron, & Luque, 2007)	3.2	-	-	2.6	34	-	-	-
AnMBR	(Zhang, et al., 2010)	7.92	-	-	5.9-19.8	58	-	-	-

*EBCT = $V_{compacted\ bed}/Q$; HRT = $V_{reactor}/Q$

*UASB: Upflow anaerobic sludge blanket, AnMBR: Anaerobic membrane bioreactor

Considering that typical activated sludge processes operate at aerobic HRT’s between 4-24 hours (Metcalf and Eddy, Inc, 2003), the CFBBR achievement of comparable nutrient removal efficiencies at considerably lower HRT’s clearly demonstrates this system’s effectiveness for biological nutrient removal.

2.3 Response to Dynamic Loading Conditions

One crucial aspect of a wastewater system is its ability to handle dynamic loadings and still treat the wastewater effectively, maintaining sufficient biological nutrient removal. There are two common forms of dynamic loading. The first is an increased flow with similar nutrient loading as before, resulting in a larger volume of diluted wastewater. An example of this would be wet weather flows (Kim & Pipes, 1996). The other form is organic shock loading, where there is a sharp increase in the organics and/or solids concentration in water, while the volume remains unchanged (Metcalf and Eddy, Inc, 2003). Both forms of dynamic loading were tested in the pilot scale system at the Adelaide Wastewater Treatment Plant in London, Canada.

2.3.1 Wet Weather Flows

Wet weather flows are a challenge for any plant in an area with frequent rain and snow. The increased volume of wastewater flowing through the same units results in a reduced residence time, thereby lowering the removal efficiency of the system. This causes the effluent to have higher than usual concentrations of pollutants. It is also possible in extreme cases, that the water must be sent through a bypass, forgoing any treatment at all. Both of these scenarios can be damaging to the environment unless handled properly (Metcalf and Eddy, Inc, 2003).

Wet weather flows were simulated in the pilot-scale CFBBR at the Adelaide Wastewater Treatment Plant. Clean tap water was added to the influent to increase the volumetric loading, thereby simulating wet weather flows. The baseline flow rate started at 5 m^3/d of degritted municipal wastewater. The clean water was then added to increase the total flow to 10 m^3/d and again to 20 m^3/d. Each of these increased flows was maintained for 4 hours, the average time for an increased wet weather flow (Chowdhury, Zhu, & Nakhla, 2010). Considering that the system was designed for a 5 m^3/d flow rate, the flows of 10 and 20 m^3/d correspond to peak flow factors of 2 and 4, respectively. A peak flow factor of 4 is a common design parameter when accounting for wet weather flows in the design a wastewater treatment system (Chowdhury, Zhu, & Nakhla, 2010).

As shown in section 4.1.2, the steady-state COD, TN, and TP removal efficiencies in the pilot were approximately 90%, 80% and 70%, respectively. During the dynamic testing, there was a measurable decrease in effluent quality and organics and nutrient removal, but still within acceptable limits. This was somewhat expected since a decrease in efficiency is the main effect hydraulic overloading (Kim & Pipes, 1996). The steady-state and dynamic effluent quality and BNR efficiencies are summarized in **Tables 2-5 & 2-6** below.

Table 2-5 Summary of Steady-State and Dynamic Loading Effluent Quality (Chowdhury et al, 2010) *(Chowdhury, Zhu, & Nakhla, 2010)*

	5 m^3/d		**10 m^3/d**		**20 m^3/d**	
Parameter (mg/L)	**Influent**	**Effluent**	**Influent***	**Effluent**	**Influent***	**Effluent**
TCOD	578	41	289	64.2	144.5	63
SCOD	192	20	96	24.5	48	22
NH_4^+-N	35.2	0.9	17.6	2.0	9.8	3.4
NO_3^--N	<0.06	5.4	<0.03	5.7	<0.2	6.9
PO_4^--P	-	<1.0	-	0.5	-	0.4
TP	12.5	1.3	6.3	1.8	3.2	2.7
TSS	443	32	221.5	-	111	38
VSS	339	22	169.5	-	85	-

*estimated from 5m^3/d influent data

Table 2-6 Summary of Dynamic Loading BNR Efficiency (Chowdhury et al, 2010) (Chowdhury, Zhu, & Nakhla, 2010)

BNR efficiency	**5 m^3/d**	**10 m^3/d**	**20 m^3/d**
COD (% removal)	90	75	49
N (% removal)	80	39	23
P (% removal)	70	43	16

The organic and nutrient removal efficiency did drop during the simulated wet weather flows. At twice the typical flowrate (10 m^3/d), the removal efficiencies and effluent quality were within acceptable parameters (U.S. EPA, 2004). However, at four times the flowrate (20 m^3/d) the removal efficiencies and effluent quality became too poor and no longer met the acceptable standards. This indicates that the maximum allowable wet weather flow at which the CFBBR could continue operating without the need for secondary clarification or chemical addition for phosphorus removal is somewhere in between, possibly about three times the baseline flow rate (15 m^3/d).

When comparing the CFBBR's response to dynamic loading with other fixed-film processes, it was found to have similar effluent quality and removal efficiency. **Table 2-7** compares the treatment efficiency of several different processes with the FBBR's performance.

Table 2-7 Comparison of Dynamic Loading effluent and nutrient removal percentages

Process	**Source**	**HRT (hrs)**	**Influent (^{1}COD, $^{2}NH_4$, ^{3}TSS, ^{4}TN) (mg/L)**	**Effluent (^{1}COD, $^{2}NH_4$, ^{3}TSS, ^{4}TN) (mg/L)**	**% Removal (^{1}COD, ^{2}TN, ^{3}TP)**
Submerged fixed-film	(Galvez, Gomez, Hontoria, & Gonzelez-Lopez, 2003)	3.2	^{1}450, ^{3}120, ^{4} 80	^{1}65, ^{2}11, ^{3}19	^{1}90%, ^{2}80%
		0.7	-	^{1}110, ^{2}55, ^{3}30	^{1}75%, ^{2}20%
Moving Bed	(Rusten, McCoy, Proctor, & Siljudalen, 1998)	1.4	^{1}527, ^{2}18.5	^{1}121, ^{2}11, ^{3}53	^{1}75%
		0.4	-	^{1}230, ^{2}18, ^{3}104	^{1}56%
Biological aerated filter	(Mann, Mendoza-Espinosa, & Stephenson, 1999)	2.0	^{1}235	^{1}57, ^{3}19	^{1}85%
		0.8	-	^{1}138, ^{3}41	^{1}35%
CFBBR	(Chowdhury, Zhu, & Nakhla, 2010)	3.2	^{1}578, ^{3}443, ^{4}61	^{1}47, ^{2}1, ^{3}31	^{1}90%, ^{2}80%, ^{3}70%
		0.8	-	^{1}65, ^{2}4.7, ^{3}50	^{1}49%, ^{2}23%, ^{3}16%

In addition to the immediate response of the system, the other important factor is the system recovery. That is to say, how fast the system returns to its steady-state effluent quality and removal efficiency. Past studies on conventional processes show that they can take anywhere from 7 to 15 days to recover from sustained peaking factors of 2.5 for 2 hours (Metcalf and Eddy, Inc, 2003). The study on the CFBBR's recovery from hydraulic overloading showed that within 24 hours after the peak flow ended, the system had nearly fully recovered. **Table 2-8** below shows the change in the attached biofilm, nitrification and denitrification rates that were measured before, during and 24 hours after the hydraulic overloading.

Table 2-8 Biomass characteristics during dynamic loading study (Chowdhury et al, 2010) *(Chowdhury, Zhu, & Nakhla, 2010)*

Parameter	**Before Overload**	**During Overload**	**24 hours after Overload**
Anoxic biofilm ($mg_{VSS}/g_{particles}$)	16.7	15.4	15.6
Aerobic biofilm ($mg_{VSS}/g_{particles}$)	6.9	6.2	6.3
Nitrification ($g_{NH4}/(g_{VSS}/d)$)	0.12	0.08	0.10
Denitrification ($g_{NO3}/(g_{VSS}/d)$)	0.34	0.28	0.31

2.3.2 Organic Shock Loading

Sharp increases in organic concentrations disrupt the biological processes occurring in the system. A large increase in biodegradable organic pollutants, without a corresponding increase in available oxygen, will result in the domination of non-nitrifying heterotrophs over the nitrifying autotrophs. This is due to the higher biomass yields and faster utilization rates of aerobic heterotrophs. This leads to washout (loss) of nitrifiers and an overall decrease of nitrification efficiency. A large loss of nitrifiers can be difficult to recover due to their slow growth rates. Ultimately, there will be an increase in effluent COD and ammonia because of the drop in efficiency of nutrient removal (Metcalf and Eddy, Inc, 2003).

In a lab scale study of the CFBBR-2 treating synthetic wastewater, the response to organic shock loading was tested. To test the response, the influent COD was increased in a step fashion. The initial influent COD concentration, starting at 420 mg/L, was increased to 720 mg/L for 4.5 hours and then increased again to 1200 mg/L for 4 hours. This corresponded to ultimate OLR of 13.2 kgCOD/m^{3}d. Liquid circulation and aeration rates remained unchanged during the shock loadings (Andalib, Nakhla, & Zhu, 2010).

During testing, the nitrification efficiencies expectedly dropped from 95% to 49% due to the heterotrophs dominating the growth, in addition to the limitations of dissolved oxygen. The DO was measured in the riser and downer as 0.0 mg/L and 2.5 mg/L at its lowest. During steady state operation, it was 0.3 mg/L and 4.9 mg/L in the riser and downer, respectively. The COD removal also dropped, although not as much, from 93.4% to 64.1%. The drop in COD removal and nitrification efficiency were seen in the effluent when both sharply increased simultaneously after 1.8 hours into the test.

Batch specific nitrification rate (SNR) tests verified the decreased nitrification efficiency, which showed a 15% decrease in activity after the 10 h carbon shock load. This indicates a 15% washout of nitrifying biomass during the shock load. Despite the changes in biomass activity, the total amount of attached biomass measured in the system did not materially change during the shock load.

2.4 Water Reuse

In addition to meeting typical secondary effluent quality standards without the need for additional treatment (clarification or chemical addition), the CFBBR has also demonstrated that it can generate an effluent that can possibly be used for non-potable reuse applications, such as agricultural irrigation or industrial uses. In order for treated wastewater to be reclaimed, it must have reasonable disinfection characteristics; meaning it can be disinfected easily without requiring large amounts of chemicals or energy. The two main requirements that generally must be met for reasonable disinfection are a BOD and TSS concentration of less than 30mg/L (U.S. EPA, 2004), as meeting these two makes disinfection through ultra-violet reasonable (Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009). The TSS is an important parameter to consider for estimating the UV dosage required for disinfection, as TSS absorbs UV radiation and can potentially shield bacteria and other microbes from the radiation; the higher the TSS concentration, the higher the required UV dose (Metcalf and Eddy, Inc, 2003). The CFBBR was capable of meeting, or closely approaching, these requirements at its steady state operation. With some additional treatment, like clarification or chemical addition, the effluent from dynamic loadings could also meet this standard.

2.5 Additional Design Considerations, Issues, and Challenges

2.5.1 Worm Predation

The solids retention time in a wastewater treatment system can have a considerable impact on the solids yield. Typically, longer SRT's lead to lower yields, since the biomass will decay to a greater extent at longer SRT's (Janssen, Rulkens, Rensink, & Van Der Roest, 1998). The solids yield can also be affected by the presence of larger organisms in the system capable of consuming the microbes, such as protozoans, metazoans, and oligochaete worms. These predators are also aided by longer SRT's, as it gives a longer time to consume the microbes (Elissen, Hendricks, Temmink, & Buisman, 2006) (Hendrickx, Elissen, & Buisman, 2010). In the past two decades, developments have been made in using worms for solids yield reduction in wastewater systems. In most cases, a separate worm reactor is employed between the activated sludge basin and the secondary clarifier (Liang, Huang, & Qian, 2006).

The effect of worm predation in the CFBBR was studied at a lab scale. In this system, the worms were active in the downer, consuming the biofilm as the particles moved down the column and when the particles moved back to the riser the biofilm would regrow until they returned to the downer to continue the cycle (Li, Nakhla, & Zhu, 2013).

Overall, it was found that simultaneous COD and nitrogen removal could be achieved with worm predation integrated into the system. The BNR efficiencies were consistent with the past studies and the system also showed greatly reduced solids yields due to the worm predation. The study revealed an observed solids yield of 0.082 gVSS/gCOD.

2.5.2 Effects of Carbon to Nitrogen Ratio on BNR efficiency

To study the effects of the carbon to nitrogen ratio, a lab scale study was conducted to examine how varying COD loading with constant nitrogen loading would affect the simultaneous COD and nitrogen removal. Since high COD concentration reduce nitrifier activity, less nitrates will be produced, and will subsequently denitrifier activity (Metcalf and Eddy, Inc, 2003). COD/N ratios of 10:1, 6:1, and 4:1 were tested at the same EBCT (0.82 hrs). The total COD removal did not vary much between the three phases, achieving above 90% removal throughout. However, the amount of COD oxidation occurring in the riser compared to the downer changed between the phases. At a COD/N ratio of 10:1, the COD oxidized in the riser was approximately 37%, due to that phase having the lowest amount of nitrates produced from nitrification. The COD/N ratio of 4:1 saw approximately 57% of the COD being oxidized in the riser, given the higher amount nitrates produced during nitrification. It was also shown that as the ratio became smaller the amount of nitrogen removal decreased, with the 10:1, 6:1, and 4:1 achieving TN removals of 91%, 82%, and 71% receptively. The first two phases both reached acceptable effluent quality (effluent concentrations: <10 mg/L TN, <20 mg/L TCOD, <15 mg/L TSS) while the third phase did not and would need additional treatment to reach an acceptable effluent quality (U.S. EPA, 2004) (Islam, Nakhla, Zhu, & Chowdhury, 2009). All three phases also showed low solid yields consistent with the other CFBBR studies, with the yields of the three phases ranging from 0.11-0.15 mgVSS/mgCOD (Islam, Nakhla, Zhu, & Chowdhury, 2009).

2.6 High Strength Wastewater Treatment

In addition to municipal wastewater, the CFBBR technology has been applied to the treatment of landfill leachate, and rendering waste. An anaerobic platform called the Anaerobic fluidized bed bioreactor (AnFBR) was applied to the treatment of wastewater sludges (primary and secondary), and thin stillage from bioethanol.

2.6.1 CFBBR

2.6.1.1 Landfill Leachate

Landfill leachate forms when organic waste in landfills is broken down by bacteria present and mixes with water in landfills, producing high concentrations of soluble COD, ammonia, and other pollutants. Treating landfill leachate effectively is of high importance due to its toxicity. The high concentrations of COD, ammonia, and heavy metals, to name just a few, can seriously damage the environment if not properly treated and removed. The low carbon-to-nitrogen ratio also makes biological treatment challenging. As the discharge limits continue to become more stringent, conventional biological treatment paired with physical and chemical treatment methods may no longer be effective enough (Haq, 2003).

The pilot CFBBR located at the Adelaide Wastewater Treatment Plant, in addition to treating municipal wastewater, was also tested for treating landfill leachate. Its integration of aerobic and anoxic conditions into one process made it a suitable candidate for achieving the higher required standards of treatment. The physical operation of this system was the same as it was for treating MWW. The anoxic riser operated in the conventional fluidization regime and aerobic downer operated in the conventional regime. Therefore, in the case leachate treatment, the CFBBR was also not run with particle circulation (Eldyasti, Chowdhury, George, & Zhu, 2010).

The pilot CFBBR was tested at various loadings and corresponding HRT's with leachate taken from the W12A landfill in London, Canada. The three flow rates used and their corresponding loading values are shown in **Table 2-9**. The average influent and effluent quality from each stage is show in **Table 2-10**.

Table 2-9 LSCFB operating conditions for leachate treatment (Eldyasti et al, 2010) *(Eldyasti, Chowdhury, George, & Zhu, 2010)*

Parameter	Column	Phase I	Phase II	Phase III
Influent (L/d)	-	650	720	864
Avg. OLR (kg COD/m^3d)	-	1.90	2.15	2.60
EBCT (d)	Aerobic	0.43	0.38	0.32
	Anoxic	0.12	0.11	0.09
HRT (d)	Aerobic	0.89	0.81	0.67
	Anoxic	0.27	0.25	0.21
SRT (d)	Aerobic	26	21	18
	Anoxic	18	17	13

Table 2-10 Influent and effluent quality of leachate (Eldyasti et al, 2010) *(Eldyasti, Chowdhury, George, & Zhu, 2010)*

Parameter (mg/L)	Influent	Phase I	Phase II	Phase III
TCOD	1259	195	197	302
SCOD	1025	149	153	245
TSS	263	56	60	58
VSS	156	38	37	44
NH_4^--N	360	34.6	35.4	54.7
NO_3^+-N	3.1	57.5	59.9	63.9
TP	6.2	1.0	1.0	1.2

The CFBBR showed considerably low solid yields. For phases I-III, the yields were 0.13, 0.15, 0.16 g VSS/g COD, which are similar to the yields seen with treating MWW in the CFBBR. In the second phase, at an OLR of 2.15 the CFBBR achieved COD, nitrogen, and phosphorus removal efficiencies of 85, 80, and 70%, respectively. These are similar removal efficiencies to CFBBR treating MWW. However, the actual effluent concentrations from the treated leachate was higher given the higher influent concentrations. The COD removal efficiencies of other treatment methods compared to the CFBBR can be seen in **Table 2-11**.

Table 2-11 Comparison of leachate treatment methods

Reactor Type	Influent COD (mg/L)	HRT (hrs)	% COD removal	Source
CFBBR	1259	8	85	(Eldyasti, Chowdhury, George, & Zhu, 2010)
Trickling Filter	800-1350	4.5	52	(Gourdon, Comel, Vermande, & Vernon, 1989)
UASB	1120-3520	24	77	(Kennedy & Lentz, 2000)
MBBR	1740-4850	36	60	(Horan, Gohar, & Hill, 1997)
FBR	1100-3800	34	82	(Suidan, Schroeder, Nath, Krishnan, & Brenner, 1993)

*UASB: Upflow anaerobic sludge blanket, MBBR: Moving bed bioreactor, FBR: Fluidized bed bioreactor

2.6.1.2 Rendering Waste

The other high strength wastewater that was treated with the CFBBR was rendering wastewater. Rendering comes from the livestock farming and food processing industry, as organic wastes are mixed together to form a high organic and nutrient concentration wastewater. Like all high strength wastewater, it must meet certain effluent quality standards before it can be discharged to municipal sewers (Del Pozo, Tas, Dulkadiroglu, Orhon, & Diez, 2003). For this study, a lab scale reactor of the CFBBR-1 configuration was built to treat wastewater from a rendering facility in Hamilton, Ontario, Canada, using lava rock as the carrier media (0.67mm diameter, 2560 kg/m^3 density). The study was carried out in three phases with varying influent flows and organic loading rates (OLR) (Islam, Chowdhury, Nakhla, & Zhu, 2009). A summary of the operating parameters of the reactor is shown below in **Table 2-12.**

Table 2-12 Summary of rendering treatment operational parameters (Islam, Chowdhury, Nakhla, & Zhu, 2009)

Parameter	**Column**	**Phase I**	**Phase II**	**Phase III**
Influent Flow (L/d)	-	2 $\pm$ 0.1	1.5 $\pm$ 0.05	1 $\pm$ 0.05
OLR (kgCOD/m^3/d)	-	14.6	11.0	7.3
HRT (h)	Anoxic	9.36	12.24	18.48
	Aerobic	39.6	52.8	79.2
EBCT (h)	Anoxic	5.52	7.36	11.04
	Aerobic	14.16	18.88	28.32
SRT (d)	Anoxic	2	4.8	20
	Aerobic	3.2	7.1	33

The CFBBR had excellent performance treating the rendering waste. In phase I, which had the highest OLR in the study, the COD removal efficiency was above 90% and the nitrogen removal efficiency was 79%. The CFBBR also had similar solids yields compared to other studies with the CFBBR, with an average yield of 0.12 gVSS/gCOD. The influent and effluent parameters of the reactor are shown in **Table 2-13**.

Table 2-13 Influent and effluent parameters of rendering treatment (Islam, Chowdhury, Nakhla, & Zhu, 2009)

Parameter	**Influent**		**Effluent**	
(mg/L)		**Phase I**	**Phase II**	**Phase III**
TCOD	29509 $\pm$ 678	3151 $\pm$ 586	2263 $\pm$ 220	1305 $\pm$ 85
SCOD	28527 $\pm$ 283	1466 $\pm$ 465	1039 $\pm$ 118	853 $\pm$ 32
NH_4-N	605.3 $\pm$ 6.2	121.8 $\pm$ 23.1	94.4 $\pm$ 9.6	0.9 $\pm$ 0.4
NO_3-N	3.8 $\pm$ 4.4	8.9 $\pm$ 2.9	5.5 $\pm$ 1.3	3.1 $\pm$ 0.7

TP	44.8 ± 5.4	34.6 ± 8.1	27.1 ± 3.3	9.8 ± 2.1
TSS	973 ± 215	2000 ± 611	1282 ± 159	461 ± 48
VSS	676 ±160	1379 ± 369	908 ± 89	330 ± 52

While the CFBBR had very high COD and nitrogen removal efficiencies, it was unable to meet sewer discharge requirements, as its effluent COD concentration was above 1000 mg/L in all phases of the study; typical sewer discharge is 300 mg/L (U.S. EPA, 2004). Many of the other parameters were also above their allowable limits for discharge However, the high removal efficiencies and low solids yields showed the CFBBR's potential for the treatment of rendering. Increasing the residence time of the rendering in the reactor could improve the treatment performance. Also, using a multistage treatment process or chemical polishing can improve the treatment performance and enable the CFBBR to meet discharge standards.

2.6.2 Anaerobic Fluidized Bed Platform

The FBBR has also been tested as an anaerobic platform for the treatment of the high strength and high solids waste streams, such as municipal sludges and corn ethanol thin stillage. A schematic of the Anaerobic Fluidized Bed Bioreactor (AnFBR) is shown in **Figure 2-9**. Like the CFBBR, the anaerobic platform utilizes a biofilm attached to a carrier media to treat the wastewater, except the microbes in this process are anaerobic. Due to this process only requiring an anaerobic environment and no other (aerobic, anoxic), a single column was used and operated in the conventional fluidization regime (Andalib, Elbeshbishy, Mustafa, & Hafez, 2014). Due to environmental requirements of the anaerobic microbes, the system also had to be maintained at 37°C and at a pH of 6.8-7.4 for ideal operation (Metcalf and Eddy, Inc, 2003).

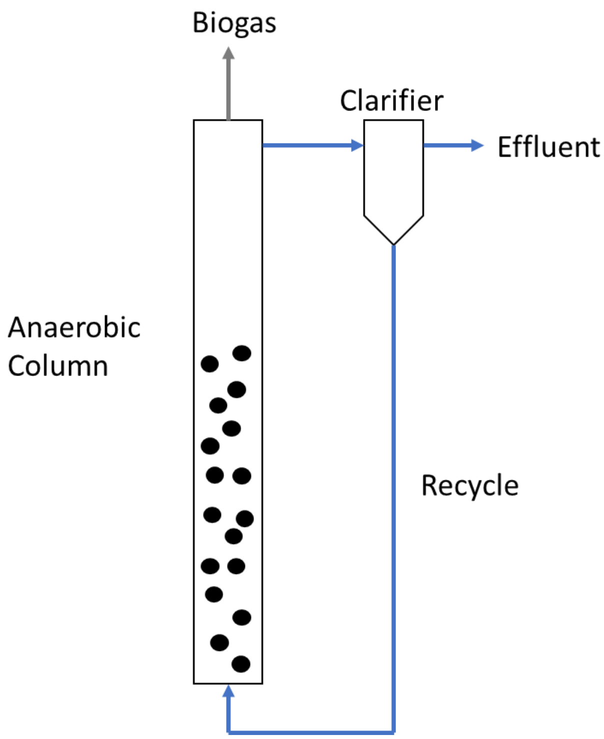

Figure 2-9 Diagram of AnFBR system (Andalib et al, 2014) *(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)*

2.6.2.1 Municipal Wastewater Sludge

Municipal sludge is a by-product of the wastewater treatment process. Primary sludge is generated from the primary clarifiers following screening and degritting of the wastewater and mostly composed of organic material. Activated sludge is settled in the secondary clarifier and later thickened to become thickened waste activated sludge (TWAS). TWAS is mostly active biomass, being composed of the bacteria and other microbes present in activated sludge. TWAS typically takes longer to treat due to it being largely composed of active biomass (Vesilind, 2003).

The AnFBR was tested for digesting both primary sludge and thickened waste activated sludge (TWAS). The digestion of each sludge was tested separately at flow rates ranging from 1.8 – 16 L/d, corresponding to HRT's ranging from 8.9-1.0 days. The average influent for TSS for primary sludge and TWAS was 38989 and 34834 mg/L, respectively, while the average TCOD was 37488 and 34414 mg/L, respectively (Andalib, Elbeshbishy, Mustafa, & Hafez, 2014). The results from treating the primary sludge and TWAS are summarized in **Tables 2-14 & 2-15** below.

Table 2-14 Summary of primary sludge treatment (Andalib et al, 2014) *(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)*

Parameter	**Phase I**	**Phase II**	**Phase III**	**Phase IV**	**Phase V**
HRT (days)	8.9	4.0	1.9	1.0	1.5
SRT (days)	17.2	6.9	2.9	1.1	1.7
VSS_{eff} (mg/L)	3693	6326	9364	21320	18069
% VSS removal	88	79	70	31	42
% COD removal	85	79	68	30	42

Table 2-15 Summary of TWAS treatment (Andalib et al, 2014) *(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)*

Parameter	**Phase I**	**Phase II**	**Phase III**	**Phase IV**
HRT (days)	8.8	4.0	1.9	2.6
SRT (days)	16.7	7.2	2.7	2.8
VSS_{eff} (mg/L)	9390	13300	20400	17800
% VSS removal	69	56	33	42
% COD removal	68	55	34	42

As expected the treatment of TWAS was less extensive than the treatment of primary sludge due to TWAS being largely active biomass and more difficult to digest, while primary sludge is predominantly inactive organic material. However, the AnFBR could effectively treat both primary sludge and TWAS at considerably shorter HRT's while maintaining comparably high SRT's due the high amount of biomass attachment (Andalib, Elbeshbishy, Mustafa, & Hafez, 2014).

The AnFBR achieved much higher VSS and COD removal efficiencies at much shorter HRT's, compared to conventional methods. It was able to achieve these efficiencies while operating at organic loading rates 5-10 times higher than conventional anaerobic digesters. The treatment results of the AnFBR are compared to several examples of conventional digestion methods below in **Table 2-16**.

Table 2-16 Comparison of AnFBR treatment capability to conventional methods

Reactor Type	Sludge Type	OLR ($kgCOD/m^3d$)	% COD removal	HRT (days)	Source
AnFBR	Primary sludge	4.2	85	8.9	(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)
CSTR	Primary sludge	2.1-2.9	33-47	10-15	(Han & Dague, 1977)
AnFBR	TWAS	4.2	68	8.8	(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)
CSTR	TWAS	1	24	20-40	(Bolzonella, Pavan, Battistoni, & Cecchi, 2005)
AnMBR	TWAS	2.4-2.6	48	7-15	(Dagnew, Pickel, Parker, & Seto, 2012)

*CSTR: Continuous stirred-tank reactor, AnMBR: Anaerobic membrane bioreactor

2.6.3 Thin Stillage

To produce ethanol as a biofuel, one of the feed stocks used is corn. The corn is mashed and fermented, producing ethanol. The leftover mash and liquid of unfermented corn is a high strength waste known as stillage that must be treated before discharge (Lee, Bae, Kim, & Chen, 2011). While stillage can be repurposed as a food source for livestock, the drying process is often not economical, due to high energy requirements. Thus, anaerobic digestion is a suitable treatment method. This way it can be converted into biogas, which can be recovered for energy production via combustion, while simultaneously removing a large portion of the organics present (Andalib, Hafez, Elbeshbishy, Nakhla, & Zhu, 2012).

The treatability of thin stillage using the AnFBR was explored. The AnFBR was fed thin stillage at an OLR of approximately 29 kg COD/(m^3/d) and an anaerobic HRT of 3.5 days, as well as a solid loading rate of approximately 10-10.8 kg TSS/m^3d. Despite the short retention time, the AnFBR achieved a TCOD removal efficiency of about 88% (Andalib, Elbeshbishy, Mustafa, & Hafez, 2014). A summary of the major influent and effluent parameters is shown in **Table 2-17**.

Table 2-17 Summary of thin stillage treatment (Andalib et al, 2014) *(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)*

Parameter (mg/L)	Influent	Effluent
TSS	46400	9800
VSS	46200	9200
TCOD	129300	14400
SCOD	62000	2700

The performance of an AnFBR relative to other anaerobic digestion technologies for the treatment of thin stillage were similar to those of treating primary sludge and TWAS. The AnFBR achieved comparable VSS and TCOD removal efficiencies at lower HRT's than conventional methods, demonstrating its great capability for treating high COD and high solids waste products. A comparison of the treatment efficiency of the AnFBR to other methods is shown below in **Table 2-18**.

Table 2-18 Comparison of AnFBR treatment of thin stillage with conventional methods

Reactor Type	OLR (kgCOD/m^3d)	HRT (days)	% COD removal	Source
AnFBR	28-30	3.5	88	(Andalib, Elbeshbishy, Mustafa, & Hafez, 2014)
CSTR	1.6-3.9	24-40	85-86	(Lee, Bae, Kim, & Chen, 2011)
ASBR	9.5	10	90	(Agler, Garcia, Lee, Schlicher,

	& Angement, 2008)

*CSTR: continuous stirred-tank reactor; ASBR: anaerobic sequencing batch reactor

2.7 Modeling

Several models have been developed for the CFBBR using the modeling programs AQUIFAS and BioWin. AQUIFAS combines activated sludge and fixed film kinetics into a single model. This model utilizes semi-empirical equations and a two-dimensional biofilm model (Sen & Randall, Improved computational model (AQUIFAS) for activated sludge, integrated fixed-film activated sludge, and moving-bed biofilm reactor systems, Part I: semi empirical model development, 2008) (Sen & Randall, 2008) (Sen & Randall, Improved computational model (AQUIFAS) for activated sludge, integrated fixed-film activated sludge, and moving-bed biofilm reactor systems, Part III: analysis and verification, 2008). BioWin models the biofilm processes as one-dimensional fully dynamic and steady-state. AQUIFAS was used for the municipal wastewater treatment modeling, while BioWin was used for modelling leachate treatment. The models were used to predict the treatment performance of the FBBR treating both MWW and leachate at various scales of operation: MWW at lab and pilot scale, and leachate at pilot scale.

2.7.1 Modeling Municipal Wastewater Treatment

AQUIFAS

AQUIFAS unifies activated sludge and fixed film processes to simulate particulate biofilm operations. It uses semi-empirical equations, incorporating Monod kinetics and mass transfer kinetics of a biofilm, to simulate biological nutrient removal. By changing the input parameters for the different loadings, the model calculates the theoretical effluent parameters and estimates the biofilm thickness on the particles. The AQUIFAS model has previously been used to successfully model the IFAS and MMBR processes, indicating its potential for modeling the FBBR process (Phillips, et al., 2008).

AQUIFAS was used to estimate the effluent parameters based on the pilot scale data (see Section 4). The error from simulated to actual results varied between 0-60%. Most simulated results were close to the actual average with deviations of 0-30%, particularly the COD, nitrogen, and phosphorus. However, the suspended solids results were off from the actual average by anywhere from 20%-67%, but still within the standard deviation (Chowdhury, Nakhla, Sen, & Zhu, 2010). The results of the simulation compared with the actual pilot study results are shown below in **Table 2-19.**

Table 2-19 Simulated vs actual data from pilot study (Chowdhury et al, 2010) *(Chowdhury, Nakhla, Sen, & Zhu, 2010)*

	Phase I		Phase II		Phase III		Phase IV	
Parameter (mg/L)	**Sim**	**Exp**	**Sim**	**Exp**	**Sim**	**Exp**	**Sim**	**Exp**
TCOD	35	26 ± 3	37	39 ± 8	45	41 ± 14	49	45 ± 7
SCOD	13	13 ± 3	9	15 ± 4	17	20 ± 8	18	23 ± 5
NH_4^+	0.8	1.2 ± 0.5	1.1	0.9 ± 0.6	1.4	0.9 ± 0.6	2.4	3.9 ± 0.9
NO_3^-	5.0	3.6 ± 1.2	5.5	4.7 ± 1.3	7.1	5.4 ± 1.3	9.9	4.8 ± 0.6
TN	7.9	6.2 ± 1.1	9.7	7.6 ± 1.3	11.5	9.4 ± 1.1	15.7	11.5 ± 1.2
PO_4^-	0.42	0.7 ± 0.1	0.34	0.5 ± 0.1	0.6	0.7 ± 0.2	0.51	0.6 ± 0.2
TP	1.12	1.0 ± 0.1	1.1	1.2 ± 0.2	1.9	1.3 ± 0.4	1.39	1.2 ± 0.4
VSS	20	11 ± 2	25	22 ± 6	25	41 ± 20	25	27 ± 6
TSS	15	9 ± 2	19	16 ± 5	17	21 ± 8	19	21 ± 6

AQUIFAS was also used to model the CFBBR-2/TFBBR process. This iteration of the model incorporated a predictive fluidization model, both two- and three-phase. The fluidization was used to link the dynamics of the fluidized bed to the biological nutrient removal efficiency more accurately. The model was based on the media type and size, flow rate and cross sectional area. It was then used to calculate parameters like bed expansion, phase hold up, and specific surface area (Andalib, Nakhla, Sen, & Zhu, 2011).

The simulated effluent data obtained was compared with experimental values from the TFBBR study and was confirmed with a two-sided t-test to be within a 95% confidence interval. The updated model incorporating fluidization, was a significant improvement over the previous AQUIFAS model. The comparative results are show in **Table 2-20.**

Table 2-20 Simulated vs actual data (Andalib et al, 2011) *(Andalib, Nakhla, Sen, & Zhu, 2011)*

Parameter (mg/L)	**Feed**	**Riser Exp**	**Riser Sim**	**Downer Exp**	**Downer Sim**
TCOD	398 ± 52	101 ± 40	97.4	50 ± 21	59.6
SCOD	118 ± 24	31 ± 8	36.1	22 ± 5	19.8
NH_4^+	30 ± 4.5	4.1 ± 0.4	4.0	0.9 ± 0.4	0.72
NO_3^-	0.8 ± 0.3	3.2 ± 1.9	3.3	5.1 ± 1.6	5.8
TP	6.5 ± 1.4	-	-	3.2 ± 0.6	6.0
TSS	214 ± 41	62 ± 30	51.2	33 ± 14	54
VSS	183 ± 30	50 ± 27	43.8	24 ± 10	37

2.7.2 Modeling Leachate Treatment

BioWin

The CFBBR was modeled with BioWin for simulated leachate treatment (see Section 6). BioWin models the CFBBR systems as 1-dimensional, fully dynamic, and steady state. It uses data on loading, biomass concentration and biofilm thickness against experimentally obtained data from large scale treatment plants. It also incorporates data on the amount of non-biodegradable and non-colloidal solids present (which are easily or readily measurable) (McGehee, et al., 2009). Since the landfill leachate had a high soluble fraction of COD, the influent specifications needed to be altered from those of typical wastewater (Eldyasti, Andalib, Hafez, Nakhla, & Zhu, 2011). The results of the simulation in comparison with the actual data from the leachate study are shown below in **Table 2-21**.

Table 2-21 Simulated vs actual data of leachate treatment in FBBR (Eldyasti et al, 2011) *(Eldyasti, Andalib, Hafez, Nakhla, & Zhu, 2011)*

		Phase I		**Phase II**	
Parameter (mg/L)	**Feed**	**Sim**	**Exp**	**Sim**	**Exp**
TCOD	1259 ± 77	236	197 ± 46	235	302 ± 98
SCOD	1025 ± 27	169	153 ± 43	169	245 ± 85

NH_4^+	360 $\pm$ 59	33.7	35.4 $\pm$ 13.1	54.7	54.7 $\pm$ 11.2
NO_3^-	3.1 $\pm$ 1.5	61.1	59.9 $\pm$ 31.1	58.4	63.9 $\pm$ 10.3
TP	6.2 $\pm$ 1.3	1.5	1.7 $\pm$ 0.3	1.8	2.0 $\pm$ 0.6
TSS	263 $\pm$ 42	60	60 $\pm$ 13	58	58 $\pm$ 8
VSS	156 $\pm$ 30	45	37 $\pm$ 5	44	44 $\pm$ 8

2.8 Bacterial Community Structure

A study on the bacterial community structure of the fluidized bed bioreactor, specifically an inverse platform using buoyant carrier particles, was conducted by Wang et al. in 2020(H. Wang et al., 2020). The main purpose of the study was to evaluate the different bacterial groups present in the aerobic and anoxic biofilms and the effluent suspended biomass and compare the results to studies on other biological nutrient removal processes.

Heterotrophs were the dominant fraction in the anoxic and aerobic biofilms, accounting for 54% and 72% by relative abundance, respectively. The fraction of denitrifiers in the aerobic and anoxic biofilm and the effluent biomass was 28%, 59% and 48%, respectively, with the dominant groups in the anoxic biofilm being *Arcobacter, Zoogloea, Thiotrix,* and *Dechlorobacter*. The significantly higher denitrifier fraction in the anoxic column is expected due the anoxic conditions favouring denitrified growth. Relative fractions of nitrifying bacteria observed were 0.56% and 0.38% for the aerobic and anoxic biofilm, respectively, with the most abundant nitrifiers observed were *Nitrosomonas* (ammonia oxidizing bacteria) and *Nitrospira* (nitrite oxidizing bacteria). While this may be much lower than expected for the nitrogen removal observed, this low abundance observed is consistent with nitrifier fractions observed in previous studies on other biological nutrient removal processes, including sequencing batch reactors, A2O and aerobic granular sludge (H. Wang et al., 2020). . The microbial community structure also indicated a significant presence of sulfate-reducing bacteria present in the anoxic zone, which contributed to the COD removal. Such insight made it possible to close the COD mass balance in the process(H. Wang et al., 2020).

2.9 Discussion

CFBBR

The CFBBR demonstrated its exceptional ability for treating municipal wastewater. It achieved COD removal efficiencies above 90%, and removal of nitrogen and phosphorus (80% and 70%, respectively) at very low hydraulic retention times. It was also able to handle higher solid loadings than conventional methods due its enhanced contact between the substrates and biofilm. Since the CFBBR was able to treat unclarified primary influent, it is possible for the influent to bypass primary clarification entirely, eliminating the need for primary clarifiers, which reduces capital costs. Overall, the CFBBR is capable of treating larger volumes of wastewater at reduced retention times than its conventional counterparts.

The longer solid retention times also lead to reduced solid/sludge yields. Low solids concentrations in the effluent stream could potentially eliminate the need for secondary clarifiers if the concentration could meet discharge standards, which in some cases it did. If not, it at least would reduce the size of the clarifiers needed and the amount of sludge produced, reducing capital and operating costs for the plant. Lower overall sludge production from wastewater treatment would also reduce the required sludge treatment capacity. Less sludge to treat would mean that smaller digesters or incinerators would be required for treatment.

High-Strength Wastewater Treatment

The single-column anaerobic platform for the CFBBR (AnFBR) had excellent results for treating high strength wastewaters. Given the CFBBR's ability to handle high solids and COD loadings, it is well-suited for treating wastes like municipal sludge and thin stillage. Conventional digesters for municipal wastewater sludge are often a large capital expenditure and require a large footprint. The significantly lower retention times of the AnFBR would allow for the same volume of sludge to treated in a much smaller reactor. This would reduce the cost and size of the digesters required to treat the sludge produced by the treatment plant. This coupled with the already lower solid/sludge yield of the CFBBR would cut down on the capital cost of treatment plants significantly.

The AnFBR is also an excellent option for treating high strength organic wastes from food industries like dairy processing plants or breweries. The AnFBR can be employed to treat the waste streams and reduce the COD and solids concentrations to meet allowable sewer discharge standards.

Modeling

The results of the modeling using both AQUIFAS and BioWin were fairly accurate with some variation of accuracy between parameters. More work will need to be done to increase the accuracy for modelling the effluent solids concentrations. However, the remaining parameters are estimated accurately. Both the AQUIFAS and BioWin models could serve as viable bases for developing future models for the CFBBR during scale-up work.

2.10 Future Perspectives

The next stage for the development of the CFBBR is scaling up to a full-scale system that can be implemented a municipal treatment plants. Since fluidization is key to the process's enhanced treatment capabilities, maintaining fluidization at a large scale will be the main focus of the scale-up work. The other aspect of scale-up will be devising methods of retaining the particles within the system or recycling the entrained particles back to the reactor. The two directions scale-up can take are developing larger fluidized beds based on the same configuration as the lab and pilot scale systems, or modifying existing system to add a fluidization component to enhance their treatment performance. When the scale-up is complete, the implementation of this system offers great potential for reducing capital and operating costs of treatment plants. The CFBBR's compact design also presents an opportunity for establishing wastewater treatment systems in a more geographically localised or isolated manner, such as on-site treatment for remote resorts or small communities with little to no wastewater piping or treatment plants. Smaller systems could also be installed for individual buildings to avoid the need for wastewater collection and piping entirely. These small, local systems would also be excellent for immediate reclamation and reuse of the wastewater instead of discharging to the environment, assuming it meets reuse standards after treatment. These options all present great potential for bringing a more effective means of wastewater treatment to places which already have established treatment systems in need of upgrades as well as remote locations which currently lack any sufficient means of wastewater treatment.

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Chapter 3

Decentralized Wastewater Treatment in an Urban Setting: A Pilot Study of the Circulating Fluidized Bed Bioreactor Treating Septic Tank Effluent

Abstract

To meet the increasing wastewater treatment demand while minimizing the land footprint of the treatment systems and plants, more efficient and compact processes are needed. The circulating fluidized bed bioreactor (CFBBR) has been proven to achieve high levels of biological nutrient removal. Past studies at the lab and pilot scale achieved 94% COD removal and 80% nitrogen removal at HRT's of 2-4 hours. A collaborative project between Western University and the Guangzhou Institute of Energy Conversion (GIEC), in Guangzhou, China, further explored the treatment of municipal wastewater with the CFBBR. A pilot CFBBR, with aerobic and anoxic columns for nitrification and denitrification, was constructed at the GIEC for in-situ treatment of septic tank effluent from a residential building. Due to high concentrations of ammonia (NH_4-N), the wastewater had a COD/N ratio of 2-3. Thus, operating at a longer HRT and supplementing COD, in the form of glucose, was necessary to achieve a high nitrogen removal efficiency. The system was run both with and without supplemental COD at HRT's between 16 and 21 hrs, treating approximately 1000-1270 L/d. Overall, a COD removal efficiency of at least 92%, ammonia removal of 97%, and nitrogen removal of 82% was achieved. The CFBBR system achieved an effluent with BOD and NH_4-N concentrations both below 5 mg/L,, a NO_3-N concentration below 15 mg/L, and a total nitrogen concentration below 25 mg/L. The compact design of this pilot-CFBBR, coupled with its high BNR performance make it an excellent option for decentralized treatment of urban wastewaters.

3.1 Introduction

Decentralized wastewater treatment (DCWWT) is becoming a more attractive method for treating wastewater for a number of reasons. One of the major challenges in wastewater treatment is the increasing volumes produced as population growth continues. This increasing demand requires either the expansion or upgrading of existing plants or constructing entirely new plants, particularly in recently developed areas. A significant issue with building new plants in developing or recently developed areas is the wastewater collection system (sewers), or lack thereof. New developments require sewers to transport wastewater to the treatment plants. However, in areas where the sewer systems are outdated or underdeveloped, connecting to the sewer may be difficult and costly (Jung, Narayanan, & Cheng, 2018). Additionally, upgrading existing plants could face difficulties because of a decreasing availability of land space due to continued urban expansion.

Rural wastewater treatment also faces its own issues. Currently, most decentralized wastewater treatment systems built for rural areas use passive or semi-passive technologies like septic tanks and lagoons. Septic tanks are typically used for areas with no sewer networks, and where homes and other buildings are widely spread out, so it would be impractical to build sewer networks. Lagoons are used in slightly larger communities that have sewer grids and enough space. Both of these technologies can have reliability issues (Oakley, Gold, & Oczkowski, 2010). Septic tanks typically discharge to septic fields, which run the risk of untreated wastewater and nutrients (nitrogen and phosphorus) leaching into the groundwater and other water bodies and causing nutrient pollution (Richards, Paterson, Withers, & Stutter, 2016; Withers, Jarvie, & Stoate, 2011; Yang, Toor, Wilson, & Williams, 2017). Lagoons are susceptible to storm flows, which can cause the flow rate to exceed the allowable tolerance and lead to discharge of untreated water into receiving waters, also causing nutrient pollution. Lagoons, if built as an earthen basin instead of concrete, also have a risk of untreated water leaching through the basin into the groundwater beneath (Oakley, Gold, & Oczkowski, 2010).

Another issue faced in wastewater treatment is the production of sludge. Sludge is produced by the removal large of suspended solids by primary clarification (primary sludge) and the removal of excess biomass from the biological treatment step via secondary clarification (secondary/waste activated sludge) (Metcalf and Eddy, Inc, 2003). Septic tanks and lagoons also produce sludge as large solids settle in the tanks/basins and must also be periodically removed. Sludge can have harmful effects if left untreated and while there many processes for sludge treatment (thickening, anaerobic digestion, chemical stabilization, incineration, etc.), they are expensive and energy intensive processes. Sludge treatment can account for 50% of a treatment plant's operating costs and 25% of the energy consumption. Furthermore, decentralized treatment plants may need to ship their sludge to an offsite sludge treatment facility (Metcalf and Eddy, Inc, 2003). These sludge treatment demands can be significantly reduced if the right biological treatment processes are chosen. Both membrane bioreactors (MBR's) and constructed wetlands (CSW’s) have shown that they are capable of treating wastewater with reduced sludge production (Capodaglio A. G., Callegari, Cecconet, & Molognoni, 2017).

The application of onsite and mechanical treatment (mechanical aeration, activated sludge, biological nutrient removal, etc.) is a possible solution for urban and rural wastewater problems. When treating urban wastewater, instead of connecting new developments to the larger sewer grid, or in cases where there are no existing sewer grids, a wastewater treatment facility along with a localized wastewater collection system can be installed (Jung, Narayanan, & Cheng, 2018). In the urban areas of developing countries, like China, where many new high-rise buildings are constructed and population density is high, decentralized systems could be a better option. Each high-rise cluster could have its own wastewater treatment system. The same could be done for housing subdivisions in suburban areas in North America. These decentralized systems can simplify the wastewater treatment process; the actual treatment systems are smaller and occupy less space, the sewer grids are easier to maintain, the need for sewage pumping stations is reduced since the sewage does not have to travel as far, or any distance at all.

For rural wastewater treatment, by using a mechanical treatment process, the risk of groundwater contamination is abated, as the system will be contained, either in a concrete or metal tank/basin. The use of mechanical treatment will also improve the stability of effluent water quality, as mechanical treatment processes can handle dynamic loads more effectively and achieve higher levels of BNR more reliably (Gill, O'Luanaigh, Johnston, Misstear, & O'Suilleabhain, 2009; Metcalf and Eddy, Inc, 2003). Furthermore, onsite treatment offers a number of opportunities for resource recovery and water reuse. Depending on the level of the treatment, the effluent could be used for a variety of purposes including irrigation or toilet flushing. Onsite treatment also offers the possibility for source separation; creating separate collection systems for grey, black, and storm water and treating each one as appropriate. This would allow for more control over water reuse and nutrient recovery (Capodaglio, 2017). However, for these new treatment paradigms to work effectively, a proper treatment process that has a small land footprint, good adaptability, and high level of treatment without producing large volumes of sludge must be used (Capodaglio, et al., 2016).

As stated above, MBR’s and CSW’s are gaining interest for these applications, however, both have problems that still hinder or prevent widespread adoption. MBR’s, though capable of achieving high nutrient removal and low sludge production, incur high capital and operating costs due to membrane capital and maintenance costs, particularly those associated with membrane fouling (Cecconet, et al., 2019). CSW's are a robust and simple process and can achieve decent treatment levels but are limited to use in warm climates because low temperatures significantly reduce their treatment capability (Capodaglio, et al., 2017).

A membrane-like biomass concentrator reactor (BCR) has been proposed as an alternative to the MBR. The BCR uses a more course filter medium for solids separation. While it is not as efficient for solid separation as the MBR, a reduction in filter fouling was observed and the BCR is still capable of achieving high levels of COD and TSS removal (Capodaglio & Callegari, Domestic wastewater treatment with a decentralized simple technology biomass concentrator reactor, 2016). A further enhancement was proposed to the BCR; the electrically enhanced BCR. By electrically charging the filter medium, fouling is further reduced and COD and TSS removal is improved as well (Cecconet, Sale, Callegari, & Capodaglio, 2019). However, the nutrient removal capabilities are still limited; 50% ammonia removal and 37% total nitrogen removal for the BCR and the nutrient removal performance of the electrically-enhance BCR has not been reported. While these technologies show promise, more study is needed on their nutrient removal capabilities before they can be adopted as alternatives for biological treatment and nutrient removal.

A technology that can be applied to solve problems in both urban and rural treatment is the Circulating Fluidized Bed Bioreactor, first developed by Jesse Zhu and George Nakhla at the University of Western Ontario (Andalib, Nakhla, & Zhu, 2010; Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009; Chowdhury, Nakhla, Zhu, & Islam, 2010; Cui, Nakhla, Zhu, & Patel, 2004; Islam, Nakhla, Zhu, & Chowdhury, 2009; Nelson, Nakhla, & Zhu, 2017). It is an attached growth biological treatment process, in which a biofilm is grown on carrier particles which are fluidized by either liquid circulation or aeration. The process utilizes two columns; an aerobic column and an anoxic column. By using these two columns treating wastewater continuously and sequentially (see **Figure 3-1)**, nitrates and nitrites generated by nitrification in the aerobic column are recirculated to the anoxic column for denitrification, achieving full biological nitrogen removal in a single process (Metcalf and Eddy, Inc, 2003).

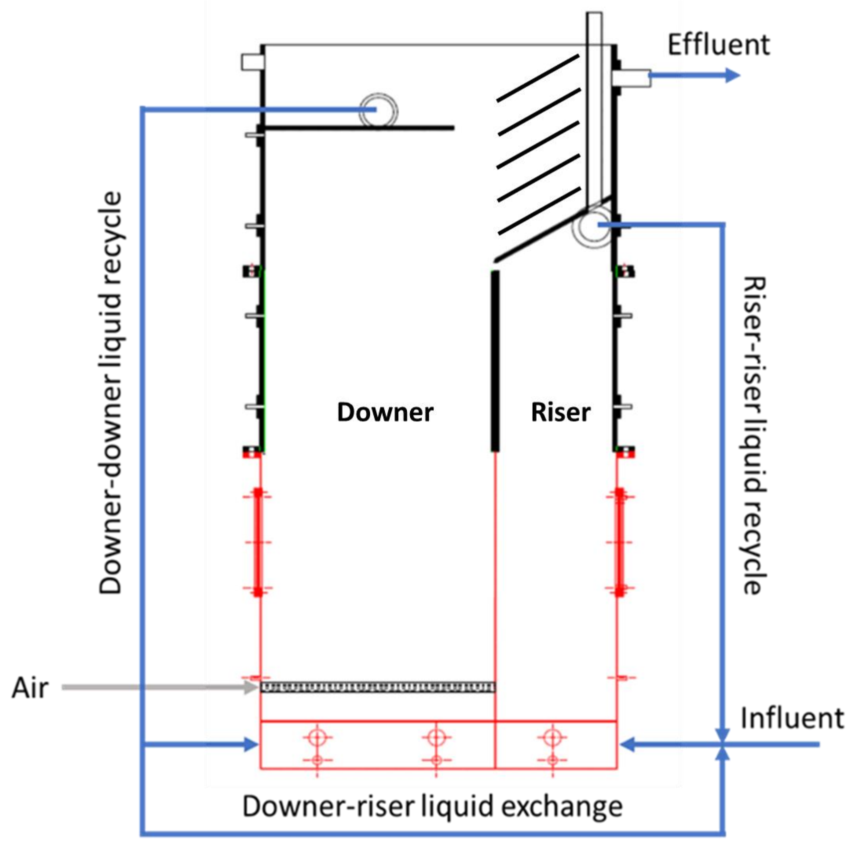

Figure 3-1 Diagram of CFBBR

A collaborative project between the University of Western Ontario and the Guangzhou Institute of Energy Conversion began to further test the CFBBR system. A pilot CFBBR was built in Guangzhou, China for the in-situ treatment of the effluent from a septic tank that takes wastewater from a residential building at the institution (GIEC). Due to the low COD/N ratio in the septic effluent (COD/N = 2-3) the system was tested at an influent of approximately 1 m^3/d with additional exogeneous COD sources to achieve a more desirable COD/N ratio (Metcalf and Eddy, Inc, 2003).

3.2 Materials and Methods

3.2.1 Apparatus

The CFBBR is a two-column process, each column containing the carrier particles. One column was equipped with an aeration and liquid circulation system (aerobic column/downer) and the other column only had liquid circulation (anoxic column/riser). The aerobic downer was 0.4 m x 0.8 m x 3 m and the anoxic riser was 0.4 m x 0.4 m x 2 m. Sampling ports were installed along the sides of the column for collecting water and particle samples. The full diagram of the CFBBR system is shown in **Figure 3-1.** The carrier particles were a polypropylene-based composite plastic (**Table 3-1**).

Table 3-1 Carrier particle properties

Parameter (units)	Value
Diameter (μm)	1387
True particle density (kg/m^3)	1328
Dry bulk density (kg/m^3)	505
Wet bulk density (kg/m^3)	1125
Minimum fluidization velocity (cm/s)	0.29
Particle terminal velocity (cm/s)	6.93

3.2.2 System Operation and Operating Conditions

The riser and downer columns both ran in the conventional fluidization regime; the riser by liquid-solid and the downer by three-phase fluidization. This was done to achieve aerobic conditions (DO > 1 mg/L) in the downer column and anoxic conditions (DO < 0.5 mg/L) to facilitate nitrification and denitrification, respectively (**Table 3-2**). To further facilitate denitrification, a liquid recycle from the downer to riser column was run to supply nitrates ($N\text{-}NO_3$) to the denitrifying bacteria in the riser. The overall flow scheme can be seen in **Figure 3-1**, with the influent entering the riser with the R-R and D-R recycles. The liquid then flows from the riser to the downer internally, where it is either recirculated in the D-D or D-R lines or exits via the final effluent port. For the first phase of operation, only the septic tank effluent was fed to the system. During the second and third phases, a solution of glucose and water was added as an exogenous COD source for denitrification. This was accomplished by a peristaltic pump, adding a 100 g/L glucose solution into the riser column along with the actual influent. The exact amount of COD supplemented was controlled by adjusting the flowrate of the glucose solution and was added in the amounts necessary to maintain a constant COD/N ratio for the influent. The liquid and air flowrates, velocities and influent flowrates are shown in **Table 3-3**. The operating parameters pertaining to retention times, loadings and biomass in the reactor are shown in **Table 3-4**.

Table 3-2 Dissolved Oxygen concentration and pH

Phase	DO-R (mg/L)	DO-D (mg/L)	pH-R	pH-D
1	0.2 ± 0.06	7.0 ± 1.7	7.4 ± 0.1	6.9 ± 0.5
2	0.2 ± 0.01	5.9 ± 1.4	7.2 ± 0.3	7.3 ± 0.2
3	0.1 ± 0.01	4.6 ± 0.6	7.0 ± 0.3	7.5 ± 0.3

Table 3-3 Flowrates, aeration and fluid velocities

Phase	Duration (days)	Q_{inf} (m_3/d)	Q_{R-R} (m3/d)	U_R (cm/s)	Q_{D-D} (m3/d)	Q_{D-R} (m3/d)	U_D (cm/s)	Q_A (m3/d)
1	28	1.2	52 ± 14	0.4 ± 0.1	168 ± 7.7	5.1 ± 0.5	0.6 ± 0.03	337 ± 155
2	70	1.02	52 ± 6.6	0.4 ± 0.04	154 ± 20	3.7 ± 0.3	0.6 ± 0.07	267 ± 38
3	83	1.27	95 ± 2.9	0.7 ± 0.02	201 ± 23	4.2 ± 0.7	0.7 ± 0.08	180 ± 23

Table 3-4 Operating Parameters

Parameter		Phase 1	2	3
Flow rate (L/d)		1200	1025	1270
HRT (h)	Riser	4.0 ± 0.2	5.5 ± 0.2	4.0 ± 1.0
	Downer	13.0 ± 0.4	15.5 ± 0.8	12.0 ± 3.3
EBCT (h)	Riser	3.3 ± 0.2	2.6 ± 0.5	2.8 ± 0.8
	Downer	6.8 ± 0.6	7.6 ± 0.5	6.3 ± 1.6
SRT (d)		93	134	105
OLR (kgCOD/m^3d)		0.38 ± 0.07	0.26 ± 0.11	0.33 ± 0.08
OLR w sup. COD (kgCOD/m^3d)		-	0.67 ± 0.33	1.29 ± 0.20
NLR (kgN/m^3d)		0.20 ± 0.01	0.11 ± 0.05	0.15 ± 0.03
Attached VSS (mg/g)	Riser	11.9 ± 5.3	11.5 ± 6.3	20.6 ± 4.2
	Downer	9.9 ± 4.1	4.4 ± 2.7	5.4 ± 1.5
COD/N (kgCOD/kgN)		2.1	2.5	2.2

COD/N w sup. COD (kgCOD/kgN)	-	6.4	8.6
Run time (days)	28	70	83

3.2.3 System Start-up

The CFBBR was started by adding the carrier particles to each column and then filling approximately half the volume with activated sludge from the Datansha Sewage Treatment Plant in Guangzhou, China. The remainder of the volume was filled with effluent from the septic tank. Aeration and liquid circulation began to induce fluidization and were continued for four days without further influent feeding to allow the microbes to begin attaching to the particles and establish a biofilm. Fluidization without influent addition was for the purpose of avoiding biomass washout before attachment could occur. After the initial attachment phase, the system was operated at 80% of the maximum influent flowrate (~800 L/d) to continue growing the biofilm.

During start-up, the primary goal was to establish a sufficient population of nitrifiers before increasing to the full influent flowrate. The specific nitrification rate of the seed sludge was tested and calculated to be 0.8 mgN-NH_4/gVSS/h. During the first four days of influent feeding during start-up the influent N-NH_4 concentration averaged 83 mg/L. The effluent ammonia peaked on day 3 at a concentration of 11.7 mg/L, after which it decreased to 3.9 mg/L on day 4 and consistently stayed below 5 mg/L for the remainder of the start-up period (see **Figure 3-2).**

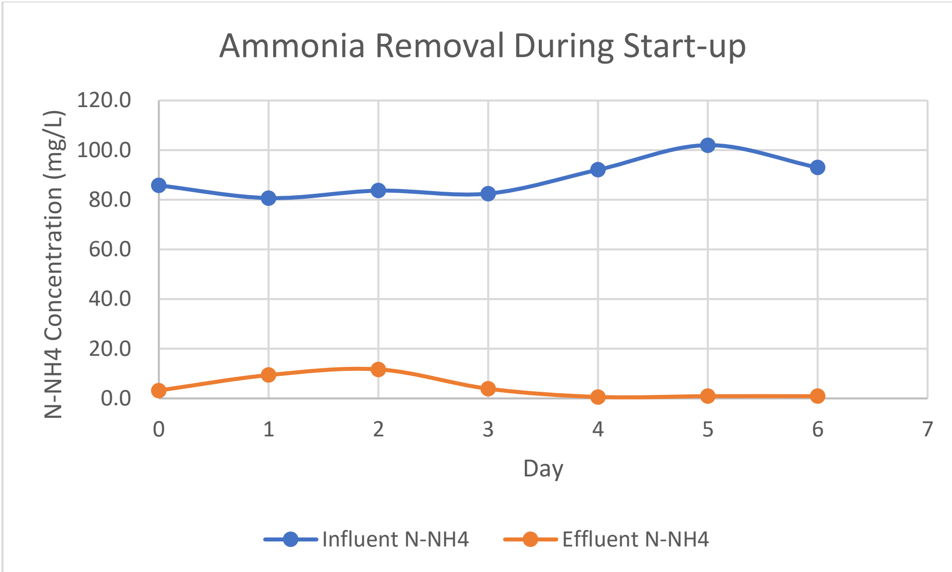

Figure 3-2 Start-up N-NH_4 concentration profile

After steady state was achieved, the influent flow rate was increased to its full operating capacity for Phases 1-3. Phase 1-3 were run for periods of 28, 70, and 83 days, respectively. Each phase was operated until steady-state operation was achieved and continued operating to collect sufficient data for steady-state analysis. Steady-state was reflected by the standard deviation of the average effluent concentration of being less than 20% of the average.

3.2.4 Analytical Methods

HACH test tube kits were used for analyzing TCOD, SCOD, TN, SN, N-NH_4, N-NO_3, and N-NO_2. TBOD and SBOD were measured using the standard 5-day dilution method. Alkalinity was measured using the acid titration method with a standard 0.02N sulfuric acid solution. The gravimetric filtration method was used for measuring TSS and VSS. Attached TSS and VSS on the particles were measured by first sonicating the particles in deionized water for 3 hours to detach the solids, and then performing the gravimetric filtration test on the water. Temperature, pH and dissolve oxygen were monitored using probes placed in the reactor.

3.3 Results

The full summary of influent and effluent characteristics is shown in **Table 3-5**, with average concentrations and standard deviations. Due to supplemental COD being used in the second and third phases, the influent COD and BOD are shown in the table as both with (eg. TCOD + C) and without the change in concentration to illustrate the initial and final characteristics of the influent. As stated in Section 2.2, the purpose of the COD addition was to maintain a certain COD/N ratio. Due to the variability in the influent COD and TN concentrations, the daily COD addition also varied, with average COD/N ratios of 6.4 and 8.6 for Phases 2 and 3, respectively. The time profiles of the influent TCOD, total nitrogen, and supplemental COD are shown in **Figure 3-3.**

Table 3-5 Summary of Influent and Effluent Characteristics

Parameter	Phase 1 (9 D.P.)		Phase 2 (7 D.P.)		Phase 3 (23 D.P.)	
	Inf	Eff	Inf	Eff	Inf	Eff
TCOD+C (mg/L)	-	-	585 $\pm$ 293	-	857 $\pm$ 133	-
TCOD (mg/L)	272 $\pm$ 50	54 $\pm$ 23	230 $\pm$ 98	45 $\pm$ 9	220 $\pm$ 32	70 $\pm$ 20
SCOD+C (mg/L)	-	-	504 $\pm$ 267	-	797 $\pm$ 129	-
SCOD (mg/L)	202 $\pm$ 42	32 $\pm$ 16	148 $\pm$ 72	30 $\pm$ 11	160 $\pm$ 22	31 $\pm$ 9
TBOD+C (mg/L)	-	-	410 $\pm$ 233	-	658 $\pm$ 77	-
TBOD (mg/L)	92 $\pm$ 11	12 $\pm$ 5	54 $\pm$ 16	7 $\pm$ 3	85 $\pm$ 19	17 $\pm$ 11
SBOD+C (mg/L)	-	-	403 $\pm$ 229	-	647 $\pm$ 80	-
SBOD (mg/L)	85 $\pm$ 11	2 $\pm$ 0.6	46 $\pm$ 15	2 $\pm$ 1	76 $\pm$ 19	3 $\pm$ 2
TN (mg/L)	139 $\pm$ 18	81 $\pm$ 18	92 $\pm$ 46	21 $\pm$ 16	100 $\pm$ 19	18 $\pm$ 10
SN (mg/L)	132 $\pm$ 16	65 $\pm$ 18	85 $\pm$ 44	19 $\pm$ 13	95 $\pm$ 20	15 $\pm$ 10
$N\text{-}NH_4$ (mg/L)	127 $\pm$ 8	25 $\pm$ 8	72 $\pm$ 36	5 $\pm$ 7	82 $\pm$ 17	2 $\pm$ 3
$N\text{-}NO_3$ (mg/L)	0.3 $\pm$ 0.2	49 $\pm$ 10	0.2 $\pm$ 0.1	13 $\pm$ 8	0.6 $\pm$ 0.3	13 $\pm$ 5
TSS (mg/L)	30 $\pm$ 4	43 $\pm$ 24	43 $\pm$ 24	17 $\pm$ 10	30 $\pm$ 6	32 $\pm$ 17
VSS (mg/L)	27 $\pm$ 4	37 $\pm$ 18	40 $\pm$ 23	14 $\pm$ 9	27 $\pm$ 5	26 $\pm$ 14
Alkalinity (mg/L as $CaCO_3$)	544 $\pm$ 51	72 $\pm$ 89*	358 $\pm$ 135	62 $\pm$ 28	486 $\pm$ 83	109 $\pm$ 45

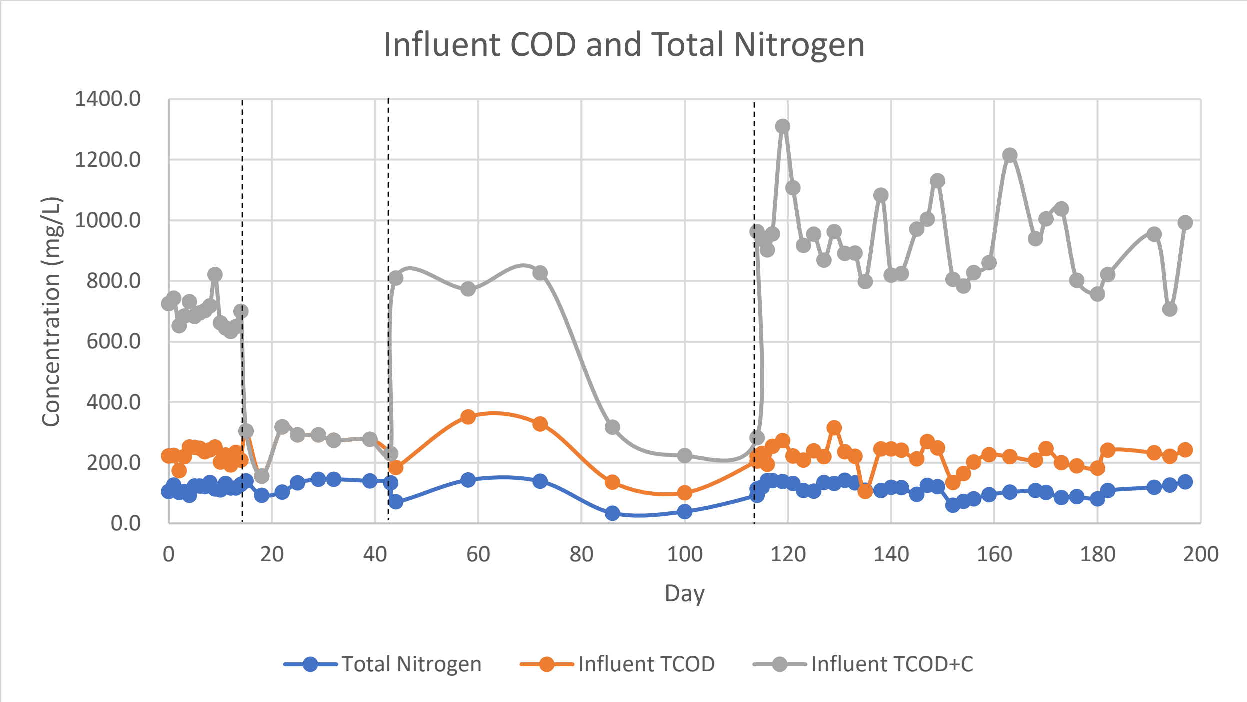

Figure 3-3 COD and TN concentration vs time profile

*dashed lines show transition points between start-up and phases 1-3

** TCOD+C denotes concentration of influent including supplemental COD

3.3.1 COD and BOD

The CFBBR had good COD and BOD removal efficiencies in all three phases. Phase 1 showed the lowest removal efficiency of COD at 80%. Phases 2 and 3 each had COD removal efficiencies of 92%. Similarly, for BOD removal, Phase 1 was the lowest, with 87% and Phases 2 and 3 had BOD removal efficiencies of 98% and 97%, receptively. The average effluent SCOD and SBOD in all three phases were close in value; ~30-32 mg/L and 2-3 mg/L, respectively. Although, the effluent TCOD and TBOD concentrations were higher than desired (45-70 mg/L COD and 7-20 mg/L BOD). This was due to the particulate COD and BOD fractions contributed by the effluent suspended solids. If it is assumed that the PCOD is removed in downstream treatment (filtration or secondary clarification), then the overall removal efficiencies increase. For COD, the removal efficiency would be 88%, 95% and 96% and BOD removal is 98%, 99% and 99% for Phase 1, 2 and 3, respectively.

3.3.2 Nitrogen

Despite the high levels of nitrogen present in the influent, the system nonetheless achieved a high amount of nitrogen removal in Phases 2 and 3. Phase 1 had significantly lower nitrogen removal, due to COD limitations for denitrification. With supplemental COD, Phases 2 and 3 achieved much higher nitrogen removal efficiencies. Total nitrogen removal for Phase 1, 2, and 3 was 42%, 77%, and 82%, respectively. Assuming removal of particulate nitrogen from downstream filtration and/or clarification, these efficiencies increase to 52%, 80%, and 85%. The effluent ammonia ($N\text{-}NH_4$) concentration in Phase 1 was 25 mg/L. Though well outside the desired effluent limits, it should be noted in the nitrogen balance (**Table 3-6)**, that Phase 1 had the highest amount of ammonia removed in terms of $gN\text{-}NH_4/d$. The final effluent $N\text{-}NH_4$ concentration was <5 mg/L for Phases 2 and 3, with $N\text{-}NH_4$ removal efficiencies of 93% and 97% for Phases 2 and 3, and effluent nitrate ($N\text{-}NO_3$) concentrations of approximately 13 mg/L in both phases.

Table 3-6 Nitrogen Balance

	Phase 1	Phase 2	Phase 3
TN_{inf} (mg/L)	138.6	92	100
$N_{synthesis}$ (mg/L)	3.1	1.5	2.8
$N_{nitrified}$ (mg/L)	104.7	85.7	95.1
Alkalinity Consumed (mg/L)	747.3	614.3	678.7
$N_{denitrified}$ (mg/L)	57.4	72.7	82.1
TN_{eff} (mg/L)	80.6	21	18
Alkalinity Recovered (mg/L)	262.1	332.1	375.0
Theoretical Net Alkalinity Consumed (mg/L)	485.1	282.2	303.7
Actual Net Alkalinity Consumed (mg/L)	471.8	296.0	377.0
Alkalinity Error %	2.8	4.7	19.4

3.3.3 Solid Retention Time and Biomass Yields

The solids retention time (SRT) of each phase was determined by calculating the amount of biomass, measured as VSS, present in the system and dividing by the mass of VSS leaving the system via the effluent (see Equation 3-1). The SRT was found to be 93, 134, and 105 days for Phases 1, 2, and 3, respectively. These long SRT values contributed to the low biomass yields, discussed below.

$$SRT\ (d) = \frac{m_{particles} * VSS_{attached}}{Q_{eff} * VSS_{eff}} \qquad (3\text{-}1)$$

The observed solids yields were determined through plotting the cumulative COD removal and VSS production and calculating the slope of the plot. The observed yield values were determined to be 0.12, 0.023, and 0.03 gVSS/gCOD for Phases 1, 2 and 3, respectively; R^2 values were 0.9867, 0.9814, and 0.9437, respectively. The observed yield values were used for the nitrogen balance calculations shown in **Table 3-6**.

3.4 Discussion

Even without downstream treatment (secondary clarification, tertiary treatment), the CFBBR already meets most of the discharge standards for the Guangzhou region, China's National Guidelines and the USEPA Secondary Discharge Limits (see **Tables 3-7 a&b**). Phases 2 and 3, with supplemental COD, met the highest Guangzhou Standards (Guangdong Province, 1990), while Phase 1, without supplemental COD, met all but the N-NH_4 standard.

Table 3-7 a & b Guangzhou Effluent Regulations and China National Effluent Guidelines

[a] Guangzhou Effluent Discharge Standards				Results		
Parameter	**Level 1**	**Level 2**	**Level 3**	**Phase 1**	**Phase 2**	**Phase 3**
TCOD	80	110	180	1	1	1
TBOD5	30	50	80	1	1	1
TSS	70	150	200	1	1	1
N-NH4	10	10	20	3	1	1

[b] China Effluent Discharge Standards					Results*		
Parameter	**Level 1**		**Level 2**	**Level 3**			
	A	**B**			**Phase 1**	**Phase 2**	**Phase 3**
TCOD (mg/L)	50	60	100	120	B (A)	A	2 (A)
$TBOD_5$ (mg/L)	10	20	30	60	B (A)	A	B (A)
TSS (mg/L)	10	20	30	50	3 (A)	B (A)	3 (A)
TN (mg/L)	15	20	-	-	-	(B)	B (A)
N-NH_4 (mg/L)	5	8	25	-	2	B (A)	A

*Level shown in ( ) represents achievable standard with secondary clarification and tertiary treatment

**USEPA 30-day average BOD and TSS secondary effluent regulations (30 mg/L) is met by Phase 2 and can be met with clarification for Phases 1 & 3

For the China National Standards (China, 2012), supplemental COD was needed for the CFBBR to meet the nitrogen-related standards, as the system was COD limited for achieving complete denitrification. The results of the CFBBR performance were evaluated without any downstream solids removal (clarification or filtration), which would not be the case in practice. If the effluent is either clarified or filtered, then Phases 2 and 3 would have met the highest standards (via removal of TSS, PCOD, and particulate nitrogen).

In order to meet the TSS and BOD_5 Secondary Discharge Standards of the USEPA, of 30 mg/L, downstream solids separation, either clarification or filtration, would have to be applied (U.S. EPA, 2004). Phase 2 already met these standards without downstream treatment. Phases 1 and 3 only exceeded the TSS standards, but by small margins; 42 mgTSS/L and 32 mgTSS/L for Phases 1 and 3, respectively. Thus, the solids removal process would be significantly smaller than a typical clarifier and less expensive than a membrane for an MBR.

Given the very low observed solids yields, the amount of supplemental COD required would be low compared to other processes, particularly suspended growth process which typically have yields of 0.3-0.4 gVSS/gCOD, as the required COD/N-NO_3 ratio would be very low for the CFBBR (**Equations 3-2 – 3-5**) (Metcalf and Eddy, Inc, 2003). The required COD/N ratio is determined by the oxygen equivalence of nitrate, 2.86 gO_2/gN-NO_3, and the COD to be oxidized. The COD_{ox} is calculated by subtracting the COD assimilated into biomass ($COD_{biomass}$) from the total COD utilized (COD_{ut}).

$$COD_{biomass} = 1.42 Y_{obs} COD_{ut} \qquad (3\text{-}2)$$

$$COD_{ox} = COD_{ut} - 1.42 Y_{obs} COD_{ut} \qquad (3\text{-}3)$$

$$2.86 NO_3 = COD_{ut} - 1.42 Y_{obs} COD_{ut} = COD_{ut}(1 - 1.42 Y_{obs}) \qquad (3\text{-}4)$$

$$\frac{gCOD}{gN-NO3} = \frac{2.86}{1-1.42*Yobs} \qquad (3\text{-}5)$$

Based on equation 3-5, a system with a biomass yield of 0.3 gVSS/gCOD would have a required COD/N ratio of 5, whereas the CFBBR, with yields as low as 0.03-0.12 gVSS/gCOD, would only require a COD/N ratio of 3.0-3.4 (**Table 3-8)**.Thus, the COD addition necessary for the CFBBR will be much lower than competing technologies, like the MBR or the MLE process (Metcalf and Eddy, Inc, 2003).

Table 3-8 Observed yield and nitrogen to COD ratio

Y_{obs} (gVSS/gCOD)	gCOD/ gN-NO_3
0.4	6.6
0.3	5.0
0.15	3.6
0.05	3.1
0.02	2.9

3.5 Full-Scale Cost and Operation Implications

A cost analysis of the COD addition and sludge processing was performed primarily using cost estimates from the USEPA (USEPA, 1985) and current prices of COD sources (glucose) and chemicals used in sludge treatment (Metcalf and Eddy, Inc, 2003). These estimates were applied to a theoretical 10 megalitre per day (MLD) treatment plant, treating wastewater of the same characteristics as the wastewater in this study (**Table 3-9)**. Two costing scenarios, one for a CFBBR-based plant and one for a suspended growth BNR (SS-BNR) plant, were devised to demonstrate the effect of COD addition and sludge cost on the economic performance of the CFBBR and SS-BNR. The influent parameters of the scenarios match that of Phase 1. The additional COD required to meet the optimal COD/N ratio was calculated using Equation 3-5 (Section 4) and the sludge production rate was determined using the assumed observed solids yields for each scenario. The yield for the CFBBR scenario was 0.12 gVSS/gCOD (Phase 1) and the yield for the SS-BNR scenario was 0.3 gVSS/gCOD (Metcalf and Eddy, Inc, 2003), shown in **Table 3-9**. Using these values, the cost of COD addition and sludge processing were estimated from costing curves from the USEPA (USEPA, 1985). The costing curves estimate cost based on annual sludge volume for each chosen treatment step.

Table 3-9 Parameters of 10 MLD Treatment Plant

Parameter	**Units**	CFBBR **C/N=Opt**	SS-BNR **C/N=Opt**	Equation/Source
Flowrate	**m3/d**	**10000**	**10000**	**-**
COD	**g/m3**	**272**	**272**	**Table 5**
	kg/d	**2720**	**2720**	**-**
TN	**g/m3**	**139**	**139**	**Table 5**
	kg/d	**1390**	**1390**	**-**
COD/N-NO3		**3.4**	**5.0**	**Eq 5**

COD suppl.	kg/yr	756,000	1,535,000	Eq 6
	$/yr	**353500**	**717500**	**Eq 7**
Yield	**gVSS/gCOD**	**0.12**	**0.3**	**Section 3.3**
Sludge	ton-ds/yr	300	1080	Eq 8
	ton-w/yr	60,000	217,000	Eq 9

3.5.1 Supplemental Carbon and Sludge Treatment Costs

The cost of the supplemental carbon was determined using the value of required COD calculated by Equation 3-5 and the typical cost of COD sources; in this case, glucose.

$$COD_{add}(\frac{kgCOD}{yr}) = \frac{COD}{N}(\frac{kgCOD}{kgN}) * TN(\frac{kgN}{yr}) - COD_{inf}(\frac{kgCOD}{yr}) \qquad (3\text{-}6)$$

$$COD_{cost}\left(\frac{\$}{yr}\right) = COD_{add}\left(\frac{kgCOD}{yr}\right) * \frac{1\ kg_{glucose}}{1.07\ kg_{COD}} * \frac{\$0.50}{kg_{glucose}} \qquad (3\text{-}7)$$

The sludge treatment train was chosen to include thickening with polymer addition, anaerobic digestion, centrifugal dewatering with polymer addition, truck hauling and landfilling as final disposal. To begin, the volume of sludge produced from biological treatment was calculated in two-steps. First, using the observed solids yield for the scenario and the amount of COD removed to determine the volatile dry solids produced and assuming 70% of the total solids are volatile. Second, the waste activated sludge (WAS) was assumed to have 0.5 percent solids to determine the wet sludge volume, as this value is needed for estimating costs from the costing curves.

$$dry\ solids\ \left(\frac{ton_{TSS}}{yr}\right) = \frac{y_{obs}\left(\frac{gVSS}{gCOD}\right) * COD_{rem}\left(\frac{ton_{COD}}{yr}\right)}{0.7(\frac{gVSS}{gTSS})} \qquad (3\text{-}8)$$

$$V_{wet}\ \left(\frac{m^3}{yr}\right) = \frac{dry\ solids(\frac{ton_{TSS}}{yr})}{0.005} \qquad (3\text{-}9)$$

The raw sludge would then proceed to chemically assisted thickening. The required mass of polymer needed was calculated based on the typical polymer to dry solids mass ratio and the cost of polymer. The final solids concentration of the thickened waste activated sludge was assumed to be 3 percent, resulting in a six-fold volume decrease of the raw WAS.

$$V_{TWAS}(\frac{m^3}{yr}) = \frac{V_{wet}}{6} \qquad (3\text{-}10)$$

Anaerobic digestion was chosen as the next step. The volume of TWAS produced by thickening was used for estimating the cost. A solids destruction of 50 percent was assumed, with a digestate solids concentration of 3 percent. This resulted in a two-fold volume decrease, post-anaerobic digestion.

$$V_{ADS}\left(\frac{m^3}{yr}\right) = \frac{0.5 * dry\ solids}{0.03} \qquad (3\text{-}11)$$

The next stage is chemically assisted dewatering of the anaerobic digester sludge. The dewatering process chosen is centrifugal dewatering with polymer addition. The polymer requirements were determined by the same method as for thickening but using only half of the original dry-solids amount, due to the 50% destruction in the anaerobic digestion stage. Therefore, the total cost of polymer for the whole treatment train is determined based on 1.5 times the total dry-solids produced and assuming 6 kg of polymer used per ton of dry solids and \$3.50 per kg of polymer (Metcalf and Eddy, Inc, 2003). The solids concentration of the sludge cake produced by dewatering was assumed to be 20 percent.

$$V_{cake}\left(\frac{m^3}{yr}\right) = \frac{0.5 * dry\ solids}{0.2} \qquad (3\text{-}12)$$

$$Cost_{polymer}\left(\frac{\$}{yr}\right) = 1.5 * dry\ solids\left(\frac{ton_{TSS}}{yr}\right) * \left(\frac{6\ kg_{polymer}}{ton_{TSS}}\right) * \left(\frac{\$3.50}{kg_{polymer}}\right) \qquad (3\text{-}13)$$

The last steps are transportation and final disposal. Transportation was assumed to be a 100-mile roundtrip and landfilling was chosen as the final disposal method, as dewatered biosolids are typically landfilled (Metcalf and Eddy, Inc, 2003). The sludge volume for both steps is the same i.e. the volume of sludge cake produced from dewatering.

The full summary of the sludge treatment process and base operating costs is shown in **Table 3-10**.

After estimating the base cost, a Construction Cost Index factor was used to account for the increase in construction costs over time. This CCI from 1985-2019 was reported to be 2.75 (RSMeans Data, 2019), so a factor of 2.75 was applied to the base cost determined from the costing curves to attain costs in \$USD 2019. Additional factors of 20% and 10% were applied for administration and laboratory costs, respectively (USEPA, 1985). The final operating costs for COD addition and sludge management are shown in **Table 3-10** and **Figure 3-4.**

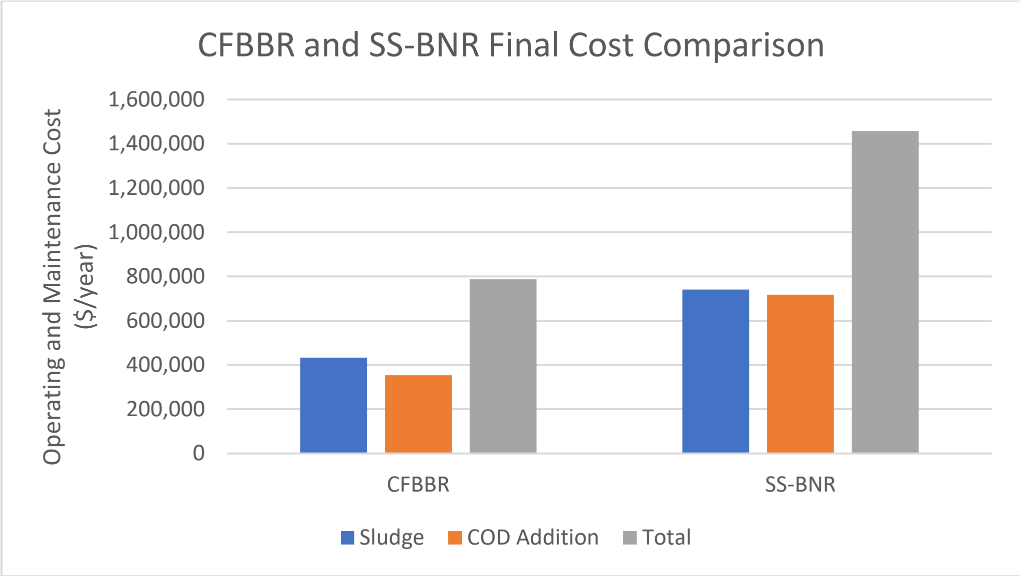

Figure 3-4 Cost analysis of CFBBR and SS-BNR Sludge Processing

Table 3-10 Base Operating Cost Estimates of Sludge Processing

Treatment Stage	Units	CFBBR	SS-BNR	Equation/Source
Thickening	V-wet (m^3/yr)	60,000	217,000	Eq 3-9
	$/yr	**16,000**	**42,000**	(USEPA, 1985)
Anaerobic Digestion	V-TWAS (m^3/yr)	10,000	36,000	Eq 3-10
	$/yr	**24,000**	**38,000**	(USEPA, 1985)
Dewatering	V-ADS (m^3/yr)	5,000	18,000	Eq 3-11
	$/yr	**25,000**	**30,000**	(USEPA, 1985)
Polymers	**kg/ton-ds**	**6**	**6**	(Metcalf and Eddy, Inc, 2003)
	$/kg	**3.5**	**3.5**	
	$/yr	**9,445**	**34,127**	**Eq 3-13**
Trucking	V cake (m^3/yr)	550	2,700	Eq 3-12
	$/yr	**20,000**	**30,000**	(USEPA, 1985)
Disposal	V haul (m^3/yr)	550	2,700	Eq 3-12
	$/yr	**37,000**	**40,000**	(USEPA, 1985)
Sum of Costs				
O&M – Sludge	**$/yr**	**477,000**	**777,500**	
O&M – COD	**$/yr**	**353,500**	**717,500**	

Total	**$/yr**	**830,500**	**1,494,500**
Unit treatment cost*	**$/kg-ds**	**1.59**	**0.72**

***Unit treatment cost = O&M – Sludge / annual dry solids production**

Given the reduced sludge processing requirements of the CFBBR compared to the SS-BNR process, the CFBBR's operating cost was the lower of the two. The significantly lower solids yield of the CFBBR is the main contributing factor. The lower yield meant that much less COD addition is required to achieve full nitrogen removal and less sludge would be produced overall. The estimated costs showed that the CFBBR's operating cost are approximately 50% of the suspended growth BNR process. It should be noted that due to the scale of economies the conventional suspended growth has a lower treatment per unit of dry-solids produced. The CFBBR has a sludge treatment cost of $1.59/kg-ds, while the conventional suspended growth process has a treatment cost of $0.72/kg-ds. However, due the significantly larger amount of solids produced by the suspended growth process, the total cost of sludge treatment was higher for the suspended growth process compared to the CFBBR.

As for the cost of COD addition, a more common and far less expensive source is sludge fermentate. Primary sludge can be fermented and the COD rich supernatant liquid can be used as COD easily attainable, inexpensive source of supplemental COD. However, this will increase the nutrient load to the mainstream process, as fermentation also releases dissolved nitrogen and phosphorus(Metcalf and Eddy, 2003).

3.5.2 Full-scale Operation

The most important challenge with scale-up of any fluidized bed process is addressing the effects of scale-up on maintaining fluidization. One of the largest three-phase processes used in industry are ebullated beds in petroleum refining, which contain a catalyst bed, liquid feed and gas bubbles (hydrogen for hydrotreating). A well-known application of the ebullated bed is the LC-Fining process by Syncrude. Modelling studies of ebullated beds show that the inner diameter of reactors can be at least 4.5m (Cheng, et al., 2014). These reactor columns were modeled to be 30+ meters tall. As such a design would not be practical for a wastewater treatment facility, a column height 5 meters would be more appropriate, as 5 meters is a typical depth for an activated sludge basin (Metcalf and Eddy, Inc, 2003).

From these dimensions, I.D. of 4.5 meters for the downer, 2.25 m I.D. for the riser and a height of 5 meters for both columns (cylindrical), the total reactor volume would be approximately 100 m^3, with liquid upflow velocity of 0.7 cm/s in the riser inclusive of a 400% internal recirculation. From this study, with an average HRT of ~18 hours, the total wastewater volume a system of this large size could treat would be ~130 m^3/day (430 person-equivalent) of high nitrogen wastewater (COD/N = 2-3). However, past laboratory and pilot studies showed that the CFBBR could treat wastewater with low nitrogen content (COD/N = 6-10) at HRT's of 2-4 hours (Chowdhury, Zhu, Nakhla, Patel, & Islam, 2009; Chowdhury, Nakhla, Zhu, & Islam, 2010). At an HRT of 4 hours, the large-scale system could potentially treat 600 m^3/day (2000 person-equivalent).

3.6 Conclusions

When evaluating its performance, ability to meet effluent guidelines, and low solids yield, the CFBBR would be a viable option for decentralized urban treatment. Its small, compact design would be best suited for heavily developed areas and its low solids yield will simplify operation as the amount the amount of sludge removal and treatment would be minimal. Additionally, it would offer more control of effluent nitrogen compared to systems like septic tanks, drainage fields and lagoons.

In terms of operational requirements, the COD addition could potentially be automated by monitoring the influent COD/N ratio and automatically adjusting the COD addition accordingly. Additionally, the CFBBR will not require as much monitoring and maintenance for sludge treatment due to its very low sludge production rate. Supplemental COD and sludge treatment cost estimates for the CFBBR were shown to be approximately 50% of suspended growth BNR systems. This is a significant factor in process selection considering the high cost of sludge treatment in existing treatment plants.

Overall, the CFBBR is an excellent option for solving the challenges facing both urban and rural wastewater treatment.

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Chapter 4

The Circulating Fluidized Bed Bioreactor as a Biological Nutrient Removal Process for Municipal Wastewater Treatment: Process Modelling and Costing Analysis

Abstract

Emerging technologies for wastewater treatment face an uphill battle to be adopted in practice because no large-scale costing data exists to prove their cost competitiveness. Similar technologies and their costing data offer some insight to the approximate cost, but more detailed estimates are required for a final decision on process selection. The circulating fluidized bed bioreactor (CFBBR) is one such technology, proven at the lab and pilot and scale, but is yet to be used on a large scale. In order to demonstrate the potential economic competitiveness of the CFBBR, a method of modifying the CapdetWorks costing software by first modeling the CFBBR in the GPS-X process simulation software was employed. The modelling was used to determine the necessary changes to a moving bed bioreactor (MBBR) process (media size, density, surface area, and bed fill fraction) in CapdetWorks to simulate the CFBBR and then generate costing estimates for both capital cost (CapEx) and operation and maintenance cost (OpEx). Benchmarking the cost estimates against simulations of conventional suspended and attached growth processes and external costing data from the US EPA was performed to both validate the costing method and analyze the CFBBR's economic competitiveness. The calculation of the net present value from the CapEx and OpEx showed that the CFBBR is predicted to have 10%-30% lower costs at low flows of 1.5 and 4.6 MGD and comparative costs to conventional processes at higher flows from 10 to 30 MGD. Furthermore, the smaller land footprint of the CFBBR-based plants and lower landfilled biosolids implies that the CFBBR's environmental footprint is superior to its competitors and offers advantages for both small-sized plants and large urban plants.

4.1 Introduction

There are numerous physical, biological and/or chemical treatment processes for removal of the different constituents of wastewater(Metcalf and Eddy, 2003). Physical treatment includes clarification and filtration, secondary treatment refers to biological treatment, and chemical treatment includes phosphorus precipitation and disinfection. The established processes are time and tested, and although new and improved technologies are constantly being developed, their adoption is hindered by the lack of reliable capital and operation and maintenance cost data.

Derivation of large-scale costs for technologies that are mostly in the embryonic development stage, i.e. laboratory testing, is very difficult. Furthermore, while piloting improves the predictability of costs, usually neither manpower nor equipment costs are linearly related to size(Sataloff et al., n.d.; US EPA, 2015; US-EPA, 1980).

Major contributors to the costs include aeration for biological treatment, chemicals for phosphorus removal, the cost of biofilm media in biofilm processes, and the treatment and disposal of biosolids produced during treatment(Metcalf and Eddy, 2003). The capital cost (CapEx) and operation and maintenance cost (OpEx) of the biological treatment and biosolids treatment processes are consistently among the highest costs for municipal wastewater treatment plants. Secondary treatment can account for 30%-40% of the total construction costs and biosolids treatment ranges from 20%-25% of the construction costs, according to the US EPA(US-EPA, 1980). According to Water Engineering: Treatment and Reuse(Metcalf and Eddy, 2003), aeration for secondary treatment accounts for 55% of all energy consumption in a typical plant on average and biosolids pumping and treatment can account for another 20% of the total energy demand. Furthermore, the inclusion of biological nutrient (nitrogen and phosphorous) removal (BNR) processes can significantly increase the capital costs of biological treatment since anaerobic and anoxic bioreactors are employed in addition to aeration tanks, as well as internal water and sludge recirculation systems.

Since costs are typically site-specific, they are developed based on local conditions, predominantly by consulting engineers and contractors, using internal software; very limited information is available in the open literature that would facilitate benchmarking of emerging technologies. Thus, the primary sources of cost data are databases and to-date very limited information is available on the use of the already sparse costing softwares such as CapdetWorks (Hydromantis) or Jacobs Engineering Group Inc.’s proprietary software, Conceptual and Parametric Engineering System (CPES). Cost and performance databases from sources like the US EPA(Sataloff et al., n.d.; US EPA, 2015) and studies that used data from existing plants(DeCarolis et al., 2012) can only give indications of costs and cost trends at different scales of well-established technologies and are unreliable and futile for costing of new technologies. This was the conclusion of a study performed by R.J. Burnside and Associates Ltd. for the Ontario Ministry of Public Infrastructure Renewal in 2005(Limited, 2005).

Statistical methods have been used to develop costing equations as well(Benedetti et al., 2006; Engin & Demir, 2006; Friedler & Pisanty, 2006; Hernandez-Sancho et al., 2011; Jung et al., 2018a; Karolinczak et al., 2020; Macal et al., n.d.; Rodríguez Miranda et al., 2015; Ruiz-Rosa et al., 2016; Singhirunnusorn & Stenstrom, 2010; Tsagarakis et al., 2003). These studies conducted statistical analyses of historical costing data and developed equations for estimating CapEx and OpEx across a wide range of flows in specific geographic locations i.e. Spain(Hernandez-Sancho et al., 2011; Ruiz-Rosa et al., 2016), Belgium(Benedetti et al., 2006), Colombia(Rodríguez Miranda et al., 2015), Poland(Karolinczak et al., 2020), Turkey(Engin & Demir, 2006), Thailand(Singhirunnusorn & Stenstrom, 2010), Greece(Tsagarakis et al., 2003), Israel(Friedler & Pisanty, 2006), and India(Jung et al., 2018b). The biological treatment processes included in the aforementioned studies were lagoons, stabilization ponds, and conventional activated sludge. However, very little data on biological nutrient removal processes was included. Furthermore, these correlations were developed for plants that met different secondary effluent criteria and provided no breakdown of costs. Given the differences in construction and labour standards, the results from the studies will require significant modifications to apply to North American standards. Just like using historical data alone, these equations only yield generalized estimates, not detailed cost estimates.

The CapdetWorks software was developed for generating detailed cost estimates of wastewater treatment plants with numerous process models for all stages of treatment(McGhee et al., 1983; Pineau et al., 1985; Wright et al., 1988). For secondary treatment this includes activated sludge, suspended and attached growth biological nitrogen and biological phosphorus removal process and several less advanced processes like lagoons and stabilization ponds. The cost breakdowns include material, chemicals, labour, and energy requirements, as well as non-construction costs like engineering design, administration, contingencies, etc, yielding detailed estimates for all plant processes. Although the software includes models for many primary, secondary, tertiary, and biosolids treatment processes, these models are not well developed to the extent of design/simulation software like GPS-X (Hydromantis ESS, Inc.), Biowin (EnviroSim Associates Ltd.), West (DHI), or Sumo (Dynamita SARL). Despite its usefulness, very few studies have been published using CapdetWorks for cost analysis of activated sludge-based or fixed-film processes and did not include any BNR processes or validate the results against any external data source(Abbasi et al., 2021; Arif et al., 2020).

Obtaining accurate and reliable cost estimates for new technologies is important for evaluating their economic competitiveness with established treatment processes. The challenge with developing reliable cost estimates for new technologies is not only the lack of cost models but also the lack of reliable process models. Thus, both technology developers and early adopters must overcome significant commercialization hurdles, without necessarily a readily available systematic approach for validation and benchmarking. The circulating fluidized bed bioreactor (CFBBR) is an attached growth process, employing biofilm grown on a fluidized carrier media (Eldyasti et al., 2010)for the treatment of degritted municipal wastewater.

The CFBBR, developed at Western University, in London, Canada achieved removal efficiencies for COD, nitrogen and phosphorus of 90%, 80%, and 65%, respectively(N. Chowdhury et al., 2009a; Patel et al., 2006) at the lab-scale. Evidence of enhanced biological phosphorus removal (EBPR) was observed and further tested; however, the low observed biomass yield reduces the potential phosphorus removal efficiency (Eldyasti et al., 2010). Since then, two pilot systems have been tested. One, designed for a maximum flow of 5 m^3/d, was designed and tested at the Adelaide Pollution Control Plant in London, Canada, demonstrating comparable performance to the previous lab studies(Eldyasti et al., 2010). The second pilot was a 25 m^3/d unit was tested in Guangzhou, China, for the treatment of septic tank effluent with a high nitrogen concentration (COD:N = 2-3). Due to the high nitrogen concentration, the system was influent flowrate was reduced to 5 m^3/d and required supplemental COD to facilitate full nitrogen removal via denitrification and demonstrated up to 80% nitrogen removal(Liu et al., 2019). An in-depth review of the CFBBR is available by Nelson et al (Nelson et al., 2017). The major highlights from this review concluded that the CFBBR can handle considerably higher loadings than conventional process as well as respond to dynamic loadings better than conventional processes, both due to the CFBBR’s high biofilm concentrations. The CFBBR platform was also shown to be adaptable for other biological treatment processes, namely anaerobic digestion of biosolids and anaerobic treatment of high-strength organic wastes, at much higher loadings than conventional technologies (Nelson et al., 2017).

A schematic of the CFBBR is shown in **Figure 4-1**. The process consists of anoxic and aerobic columns, with internal liquid circulation for each column to maintain fluidization. Nitrification and aerobic BOD removal occur in the aerobic column and denitrification occurs in the anoxic column. Influent enters at the base of the riser, flows upward and transfers to the bottom of the aerobic column where it flows upward and is recycled to the anoxic column, and the final effluent leaves at the top of aerobic column. Liquid is circulated between the columns to transfer nitrates from the aerobic column to the anoxic column for denitrification, with the remaining BOD and ammonia transferred from the anoxic to the aerobic columns to facilitate organics removal and nitrification, respectively. Simultaneous nitrification denitrification (SND) has been demonstrated to occur in the aerobic column(Islam et al. 2009) .

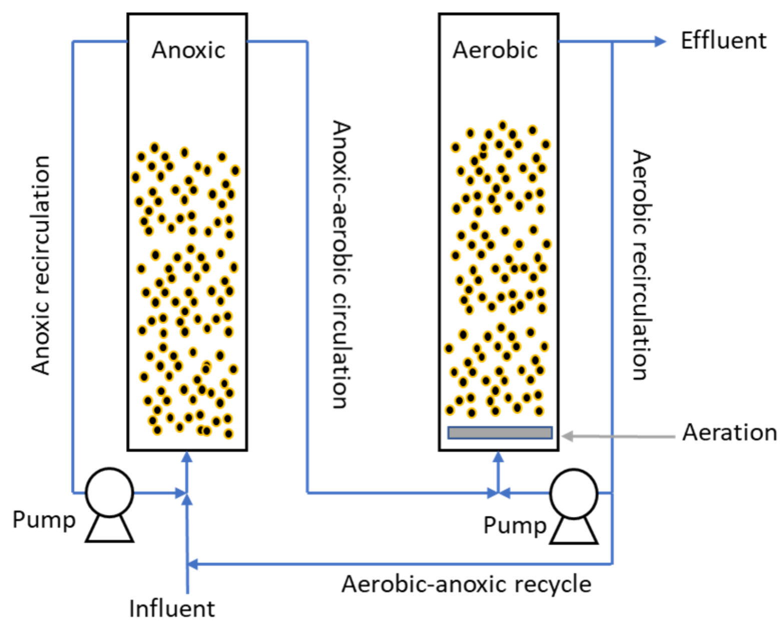

Figure 4-1 Schematic of the circulating fluidized bed bioreactor

The purpose of this study is to evaluate the economic competitiveness of the CFBBR against conventional suspended and attached growth biological nutrient removal technologies. In this work, we outline a hybrid approach that though used specifically for the CFBBR can be used for other emerging technologies. The sequential approach involves system design using GPS-X (or Biowin, SUMO, etc.), followed by incorporation of process sizes in the CapdetWorks, which allows for design overrides to the process units. Rather than developing a new process model for the technology, which is not readily incorporated in commercial simulation/design software, the approach involves detailed calibration of an existing unit process based on lab/pilot data, as an alternative to modelling the CFBBR, and subsequent use of the calibrated model for full-scale design. A critical component of the procedure involves benchmarking the costs of competing technologies with data from existing databases, such as the US EPA, to ensure the reliability of the CapdetWorks cost predictions. While the limitations of this approach with respect to the accuracy of cost estimates in light of the lack of site-specific conditions are definitely pertinent, for benchmarking of emerging technologies, the relative costing is more important. Currently, the CapdetWorks software does not have a unit for the CFBBR. However, the MBBR is somewhat similar in construction and operation to the CFBBR(Metcalf and Eddy, 2003). Therefore, modifications of the dimensions of the MBBR as well as the biofilm media can give a close approximation. However, these modifications may lead to inaccuracies in the prediction of performance and effluent quality of the costing simulations in CapdetWorks.

4.2 Modelling and Simulation Methods

To overcome the challenge of modifying MBBR units in CapdetWorks to simulate the CFBBR and to ensure that the conventional processes being simulated all achieve comparable treatment, they were first modelled in GPS-X for multiple flows ranging from 1.5 to 30 MGD. The GPS-X model was calibrated with the results of Liu et al. (Liu et al., 2019). This study detailed the operation of a CFBBR treating approximately 1.2 m^3/d of low COD/N septic tank effluent with supplemental carbon provided for denitrification; the operating HRT was approximately 18 hrs. The model yielded effluent predictions all within the standard deviations of the experimental results for all TSS/VSS, COD/BOD, and nitrogen parameters. The detailed effluent predictions are shown in the supplemental information in **Appendix 1 - Table A1.1**.

The suspended growth processes chosen were the Modified Ludzac-Ettinger (MLE) process, 4-Stage BardenPho (4-BDP), A2O, UCT, and 5-Stage BardenPho (5-BDP). For biofilm processes, the biological aerated filter (BAF), moving bed bioreactor (MBBR), and integrated fixed film activated sludge (IFAS) processes were chosen and paired with a pre-anoxic stage to facilitate full nitrogen removal via pre-denitrification(Metcalf and Eddy, 2003). Each of these processes were modelled for the treatment of a medium strength municipal wastewater (COD = 430 mg/L, BOD 220 = mg/L, TSS = 220 mg/L, VSS = 170 mg/L, TKN = 40 mg/L, N-NH_4 = 25 mg/L, TP = 10 mg/L, SP = 8 mg/L, Alkalinity = 250 mg/L as $CaCO_3$) at 1.5, 4.6, 10, 20 and 30 MGD flows.

30 MGD was chosen as the maximum flow size because a 20-module CFBBR plant can handle a flow of 30.4 MGD operating at a total HRT of 3 hours(N. Chowdhury et al., 2008).

The full plants simulated included pump stations, preliminary treatment, primary and secondary clarification, chemical phosphorus removal, and disinfection for the liquid train and gravity thickening, anaerobic digestion, centrifugation, and biosolids hauling and landfill for the biosolids treatment train. All enhanced biological phosphorous removal (EBPR) processes employed chemical phosphorus precipitation after anaerobic digestion in the biosolids train. The CFBBR did not use primary clarification as previous studies have demonstrated that it is capable of treating primary influent even at low HRT's(Andalib et al., 2010b; N. Chowdhury et al., 2008; Nelson et al., 2017). The main design parameters for each process are listed below in **Table 4-1**, derived from literature sources(Metcalf and Eddy, 2003)(Nelson et al., 2017). All processes were designed to achieve an effluent Total Phosphorus <1 mg/L and at least 80% total nitrogen removal. Examples of the process flow diagrams (PFD) in CapdetWorks are presented in **Figure 4-2a – 4-2c**.

Table 4-1a Major design inputs for modelled processes

Parameter	Stage	MLE	4BDP	A2O	UCT	5BDP	IFAS	BAF	MBBR	CFBBR
HRT (hr)	Anaerobic	-	-	1	1.5	1	-	-	-	-
	Anoxic 1	2.5	2.5	2	3	2.5	2	1	2	1
	Aerobic 1	12	14	10	8	14	8.4	2	8	2
	Anoxic 2	-	1.66	-	-	1.66	-	-	-	-
	Aerobic 2	-	0.33	-	-	0.33	-	-	-	-
Recycle (-)	Ae-Ao Recycle	4	4	3	3	3	3	3	3	4
	Ao-An Recycle	-	-	-	2	-	-	-	-	-
	RAS Ratio	0.6	0.6	0.75	1	0.75	0.6	-	-	-
Bed fill (%)		-	-	-	-	-	50	50	50	50
Ferric Dose (kg/kg)	Chem P	4	4	4	4	4	4	4	4	4
Overflow Rate ($m^3/m^2/d$)	PC	40	40	40	40	40	40	40	40	40
	SC	20	20	20	20	20	20	20	20	20
Solids Loading Rate ($kg_{TSS}/m^2/d$)	GT	9	9	9	9	9	9	9	9	9
Solids Loading Rate ($kg_{TSS}/m^3/d$)	AD	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5

Table 4-1b CFBBR Modular Design; 1.5 MGD per module

Design Parameter	Anoxic Column	Aerobic Column	Units
Column Height	30	30	m
Column Diameter	3.2	4.5	m
Superficial Liquid Velocity	0.01	0.01	m/s
Aerobic Recycle Ratio	-	4	-
Air Flow Rate	-	0.8	m^3/s
HRT	1	2	hr
Bed Fill	50	50	%
Biomass Yield	0.18	0.18	gVSS/gCOD

Table 4-1c CFBBR Module Flow Sizing

Module Count	Anoxic Volume (m^3)	Aerobic Volume (m^3)	Total Volume (m^3)	Flow Rate @ 3 hr HRT (MGD)
1	240	480	720	1.5
3	720	1440	2160	4.6
7	1680	3360	5040	10.7
13	3120	6240	9360	19.8
20	4800	9600	14400	30.4

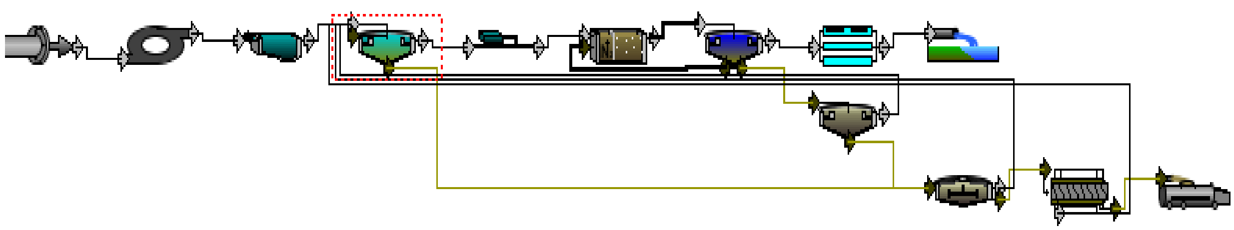

Figure 4-2a MLE Plant PFD

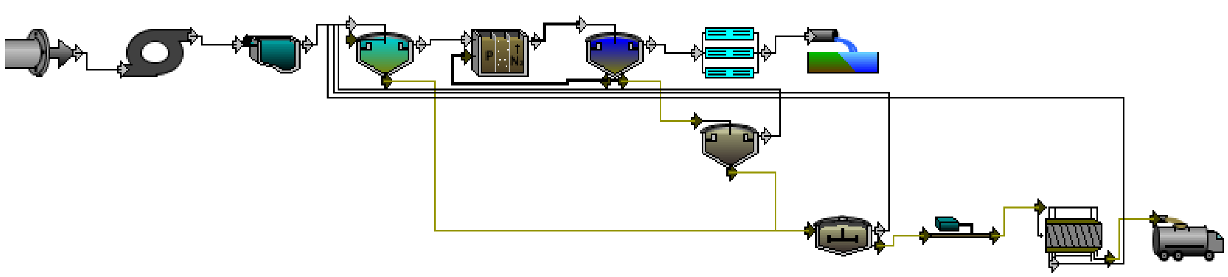

Figure 4-3b A2O/UCT/5-BDP Plant PFD

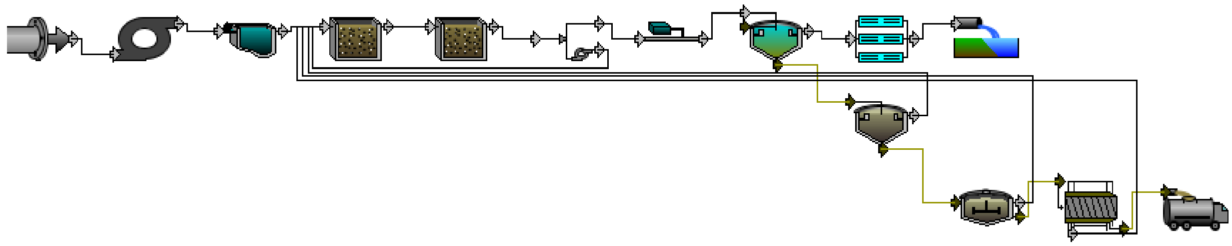

Figure 4-2c CFBBR Plant PFD

Past lab and pilot studies were reviewed to determine the appropriate sizing for the CFBBR units, as well as the biomass yield, liquid pumping requirements for fluidization. These simulations assume the CFBBR will be a conventional fluidized bed, meaning that the carrier media will be more dense than water and fluidized by an upwards flow of air and water(Andalib et al., 2010b; N. Chowdhury et al., 2008; Nelson et al., 2017). Based on these studies, a total HRT of 3 hours was chosen for the CFBBR and the biomass yield in both the GPS-X and CapdetWorks simulations was 0.18 gVSS/gCOD.

The fluidization energy required was calculated using **Equation 4-1** for liquid pumping energy,

$$E = Q * \rho g h * 0.7 \qquad (4\text{-}1)$$

Where Q is the wastewater flowrate, ρ is the density of water, g is acceleration due to gravity, and h is the depth of the CFBBR unit. The flow rate is calculated by assuming a superficial liquid velocity of 0.01 m/s to maintain fluidization. This liquid velocity is based on the combined influent and liquid recirculation flows. The pump efficiency was assumed to be 70%(Metcalf and Eddy, 2003).

The CapdetWorks results were compared against two estimating methods from the US EPA. The total cost of the BNR systems were verified against a costing report published by the US EPA in 2007(Sataloff et al., n.d.). This report included process CapEx for numerous BNR processes across multiple scales. The cost estimates for the sludge train were verified by using a cost estimation method published by the US EPA in 1985; most recent available source on cost estimates for sludge and biosolids handling(EPA, 1985). The calculated sludge production for each simulated plant was applied to costing curves to get a base capital value, indexed for 2021 $ using the ENR index, and cost multipliers were applied for line items such as engineering design, administration, legal, and contingencies.

Net present values (NPV) were determined using interest rates of 3%, 5%, and 7% and a thirty-year lifespan for the plant, using **Equation 4-2**.

$$NPV = P + R\frac{(1+i)^n-1}{i(1+i)^n} \qquad (4\text{-}2)$$

Where P and R are the CapEx and OpEx, respectively, *i* is interest, and n is the life span of the plant.

4.3 Results and Discussion

4.3.1 GPS-X Modelling

To ensure that all cost estimates for the processes in this study are comparable, all 9 technologies were first modelled and simulated in GPS-X. The final effluent achieved for each process is shown in **Table 4-2**. All technologies were designed to meet a minimum TN removal efficiency of 80%, and an effluent TP of < 1 mg/L. The design parameters used in these simulations were then applied to CapdetWorks to produce the cost estimates for each process and plant simulated. From the effluent results in **Table 4-2,** it is apparent that all processes evaluated in this study, including the CFBBR, can achieve effluent quality far superior to the standard secondary effluent discharge limits set by the US EPA; <30 mg/L TSS and BOD_5. While tertiary standards are site-specific, effluent BODs, total nitrogen, and total phosphorous of 1.5-5, 5.2-9, and 0.1-0.5 mg/L respectively are typical of tertiary effluents.

Table 4-2 Influent and final effluent characteristics from GPS-X

Parameter		Influent	Effluent								
			MLE	4-BDP	A2O	UCT	5-BDP	IFAS	MBBR	BAF	CFBBR
TCOD	mg/L	430	24	25	35	36	35	25	21	16	30
SCOD	mg/L	148	17	17	24	25	24	18	19	15	21
scBOD5	mg/L	90	2.2	2.5	2.1	2.3	1.8	2	2.8	1.4	1
cBOD5	mg/L	220	4.2	4.2	5.1	5.5	5.2	3.4	3.3	1.5	3.1
TSS	mg/L	220	15	15	14	15	14	15	5	0.6	16
VSS	mg/L	170	4.9	4.8	6.7	6.9	7.1	4.2	1.6	0.2	6.3
TKN	mg/L	40	2.1	1.9	2.7	2.8	2.5	2.1	1.6	1.4	1.1
N-NH4	mg/L	25	0.3	0.2	0.4	0.4	0.1	0.4	0.6	0.1	0.2
N-NO3	mg/L	0	6	2.5	6.2	5.6	2.6	5	3	4.7	5.6

N-NO2	mg/L	0	0.3	0.3	0.3	0.4	0.2	0.4	0.6	0.1	0.2
TN	mg/L	40	8.4	4.7	9.3	8.9	4.3	7.5	5.2	6.3	7.1
TP	mg/L	10	0.1	0.1	0.2	0.2	0.2	0.5	0.5	0.2	0.5
Alk	mg/L	250	76	83	77	78	88	76	85	75	61
DO	mg/L	0	2	2	2	2	2	2	2	2	2
pH	mg/L	7	7	7	7	7	7	7	7	7	7

It is also observed that all processes, including the CFBBR, achieved TN below 10 mg/L and TP below 0.5 mg/L and the 4-BPD and 5-BDP process achieved TN below 5 mg/L. The attached growth processes, particularly the BAF and MBBR, had much lower effluent TSS and VSS compared to the suspended growth processes. Clearly it can be seen that the BDP processes provide the best nitrogen removal and attached growth processes provide lower effluent solids. Though the CFBBR did not show any particular advantage in terms of effluent quality, it must be noted that this effluent quality was achieved at a far lower operating hydraulic retention time (HRT) compared to the other processes or conversely in a much smaller process unit than other processes.

4.3.2 Cost Estimates Verification

To verify the results, the estimated BNR system capital costs were compared to the reported values from the US EPA(Sataloff et al., n.d.). The largest variability shown by US EPA data was the capital cost of BNR systems at flows below 5 MGD where costs varied from approximately \$1/GPD to \$9/GPD. Both the higher cost per gallon of flow and greater variability in cost at lower flows is to be expected. Site-specific factors as well as non-construction costs have a greater effect on the costs at lower flow sizes due to the economies of scale. At higher flows, the variability in cost becomes less pronounced, ranging from approximately 0.5 \$/GPD to 2.3 \$/GPD (Caldas et al., 2019; Hernández-Chover et al., 2018). As seen in **Figure 4-3a**, the CapdetWorks results compare well with the US EPA, particularly at flows above 5 MGD.

The biosolids management costs were benchmarked against estimates derived from a US EPA estimating method. The calculated biosolids volumes and concentrations from CapdetWorks were applied to costing curves for gravity thickening, anaerobic digestion, centrifugal dewatering, and hauling and landfilling (EPA, 1985) to get the base capital costs. These estimates were then indexed for 2021 $ using the ENR index (ENR 1985-2021 = 2.694). Finally, cost multipliers for the non-construction line items engineering design (10%), supervision (5%), legal and administration (2%), and contingencies (10%) were applied to derive the final CapEx estimate. **Figure 4-3b** plots the US EPA indexed estimates against the cost values predicted by CapdetWorks. The US EPA-produced estimates were 92% of CapdetWorks' estimates, i.e., the discrepancies between the two methods were 8%. It must be noted that the available costing curves are limited to a maximum annual biosolids volume of one million gallons. As a result, the biosolids treatment CapEx could only be estimated for influent flows of 1.5, 4.6 and 10 MGD.

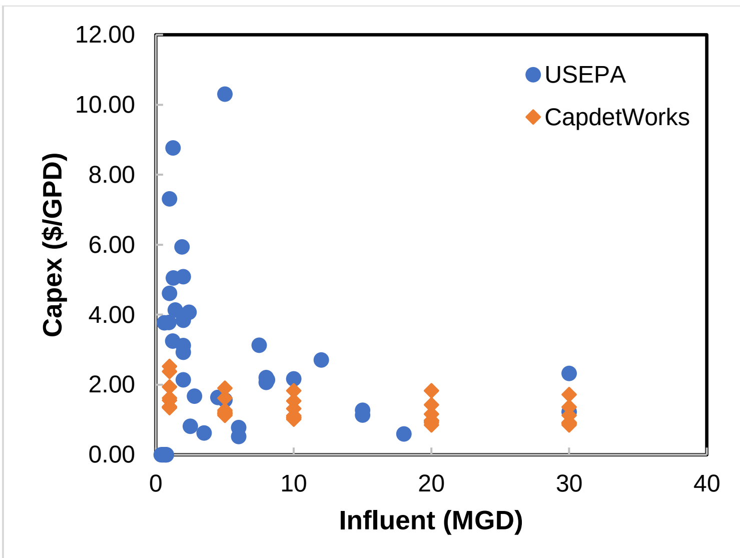

Figure 4-3a Secondary Treatment CapEx estimates of US EPA and CapdetWorks

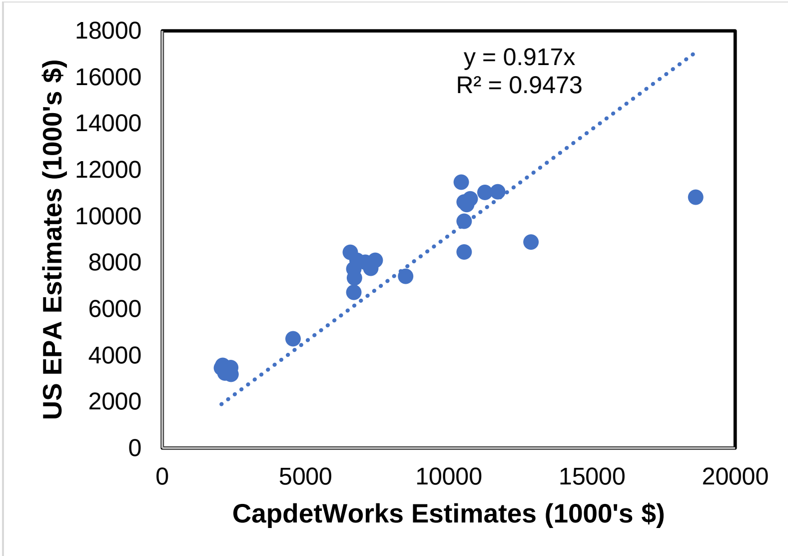

Figure 4-3b US EPA and CapdetWorks Sludge Processing CapEx estimates

By verifying the results from CapdetWorks against two different methods from the US EPA, historical data (Sataloff et al., n.d.) and costing curves (EPA, 1985), it was concluded that the CapdetWorks model produces reliable results across multiple scales.

4.3.3 CapEx

CapEx and OpEx estimates for all processes are shown in **Appendix 1 - Figures A1.1 – A1.10**. The units of cost on the y-axis are thousand $. For the CFBBR OpEx estimates, the "BNR Total" includes the energy required for fluidization of the carrier media. Unlike the aeration energy, which could be estimated directly from CapdetWorks, the liquid pump energy was calculated separately using **Equation 4-1** and added to the final estimates. All unit costs ($/kWh, $/hr-labour, $/acre-land, etc.) are tabulated in **Appendix 1 - Table A1.12**.

The CapEx ratios of the CFBBR to the average of the suspended and attached growth process are graphed in **Figure 4-4a.** The most striking finding inferred from **Figure 4-4a** and **Figures A1.1 – A1.5** is that the CFBBR has considerably lower CapEx estimates at flows of 1.5 to 30. CapdetWorks showed that the largest cost savings are due to the lack of primary clarifiers, the CFBBR's low unit costs, and reduced biosolids production. Overall, the CFBBR was estimated to achieve 30% - 45% reduction in CapEx compared to the conventional processes. The elimination of primary clarification in particular, as well as the reduction in size for secondary treatment and biosolids treatment units means that a CFBBR-based plant will occupy significantly less land space; a very important consideration for urban treatment plants and plants with limited space. CapdetWorks estimates footprint as the larger of two calculations: the area based on flow and total area required by all treatment processes. Thus for the CFBBR, the area based on flow, which is larger than the process area, governs and hence land costs for the CFBBR are overestimated. However, as the total land cost is only 1%-3% of the CapEx for all simulated processes and flows, this overestimation will have little effect on the CapEx and NPV of the CFBBR, but still remains an important practical consideration for the process selection of space limited treatment plants.

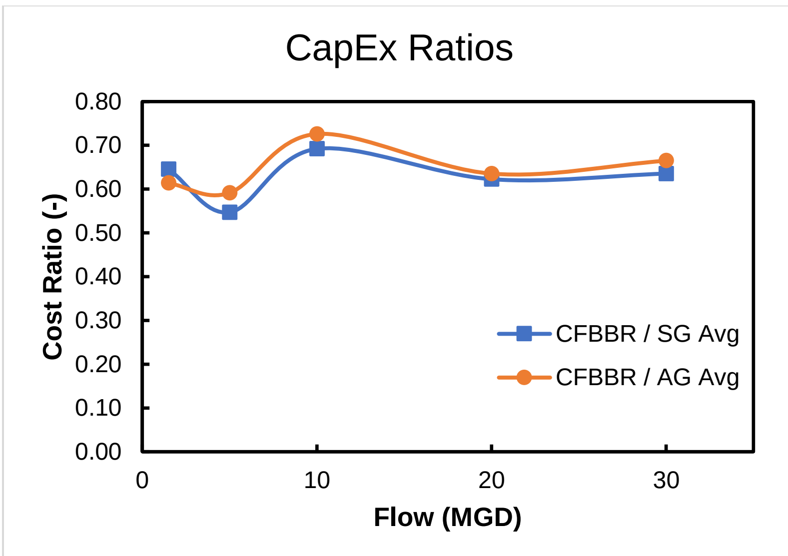

Figure 4-4a CFBBR / Conventional CapEx ratios at all scales

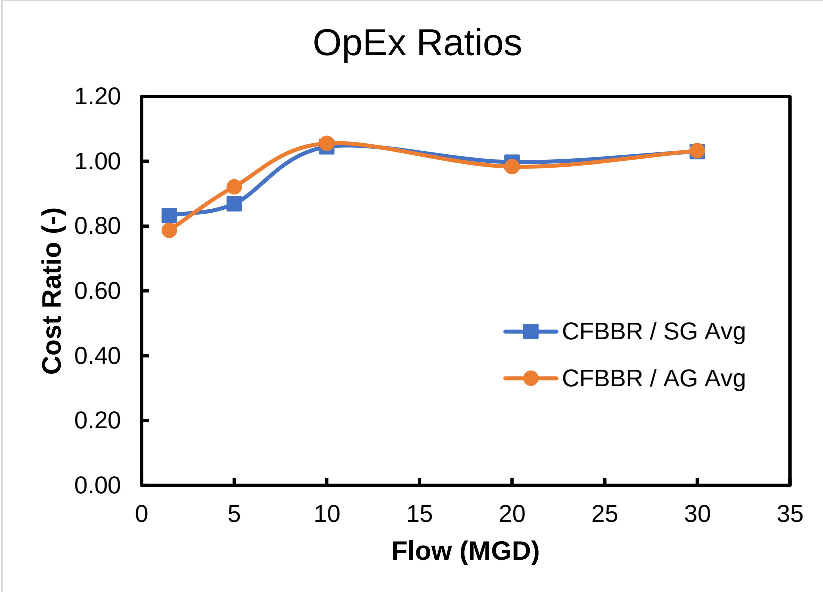

Figure 4-4b CFBBR / Conventional OpEx ratios at all scales

The CFBBR's ability to forgo primary clarification before secondary treatment offers CapEx reduction since primary clarification accounted for 3%-6% of the total CapEx for the conventional suspended and attached growth processes from 1.5 MGD to 30 MGD flow sizes. Forgoing primary clarification also led to a reduction in overall biosolids treatment OpEx, to be discussed in the *OpEx* section later.

The enhanced treatment capabilities of the CFBBR also enable it to operate at considerably lower HRT's and handle higher organic loadings compared to conventional processes(N. Chowdhury et al., 2008). These lower reactor volume requirements contribute to the CFBBR's significantly lower CapEx. This advantage becomes less pronounced at higher flows due to the size limitations of fluidized beds. The higher the flow, the more modules required and thus greater material costs, regardless of the operating HRT. This is observed in **Figure 4-4a-b**, which plots the ratios of the CFBBR CapEx and OpEx to the suspended and attached growth systems' average CapEx and OpEx, respectively. The lowest ratios for both the suspended and attached growth averages (0.55 and 0.59, respectively) are at 4.5 MGD, where the CFBBR has the largest CapEx savings. The ratios increase at 10 MGD (0.69 and 0.73) and maintain similar levels at 20 and 30 MGD.

Figures 4-5a-b plot the CapEx ratios of preliminary & primary, secondary and biosolids treatment by flow size. From **Figure 4-5b**, it can be seen that only the secondary treatment system CapEx ratios show a sharp change between from 4.6 to 10 MGD, while primary and secondary CapEx ratios show no sharp changes between flow sizes. This shows the CFBBR's diminished cost effectiveness at higher scales, which was due to the CFBBR's high module count. **Figure 4-5b** shows the CapEx ratios for the secondary tanks only, excluding the cost of blowers and chemical feed systems. The lowest CapEx ratio is at 1.5 MGD and increases at the higher flows before leveling above 10 MGD. Combined, **Figures 4-5a-b** indicate that the reduction in tank costs for the CFBBR is considerable but is partially offset by increased aeration requirements, i.e., the blowers costs. The cost of blowers is an important contributor to the CapEx due the CFBBR's higher aeration requirements. The CFBBR's blowers cost 40%-50% more than its competitors due to the higher loading processed by the CFBBR, as a result of elimination of primary clarification, and subsequent higher oxygen requirements.

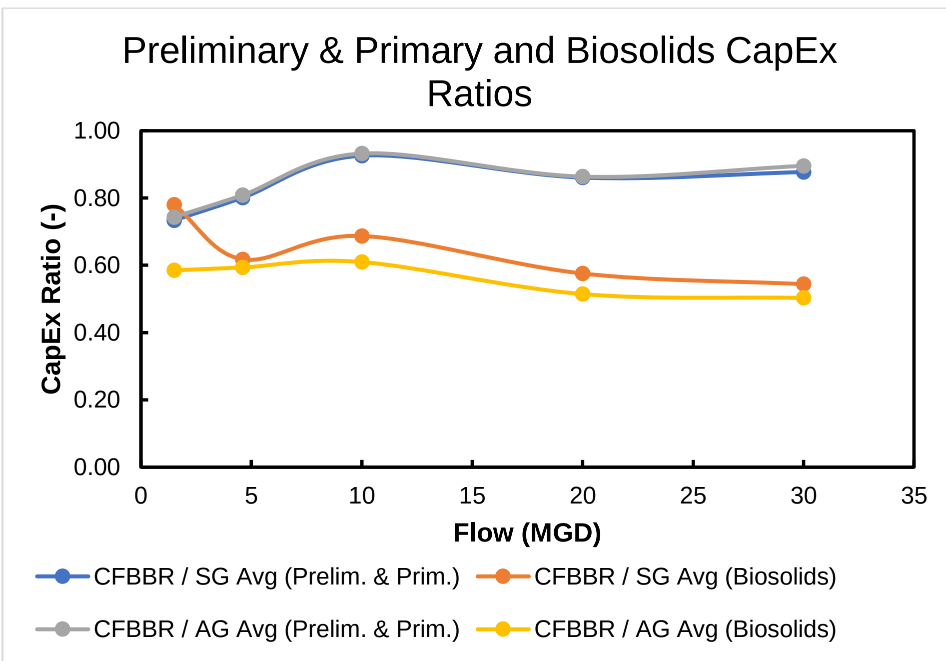

Figure 4-5a CapEx ratios for preliminary and primary treatment

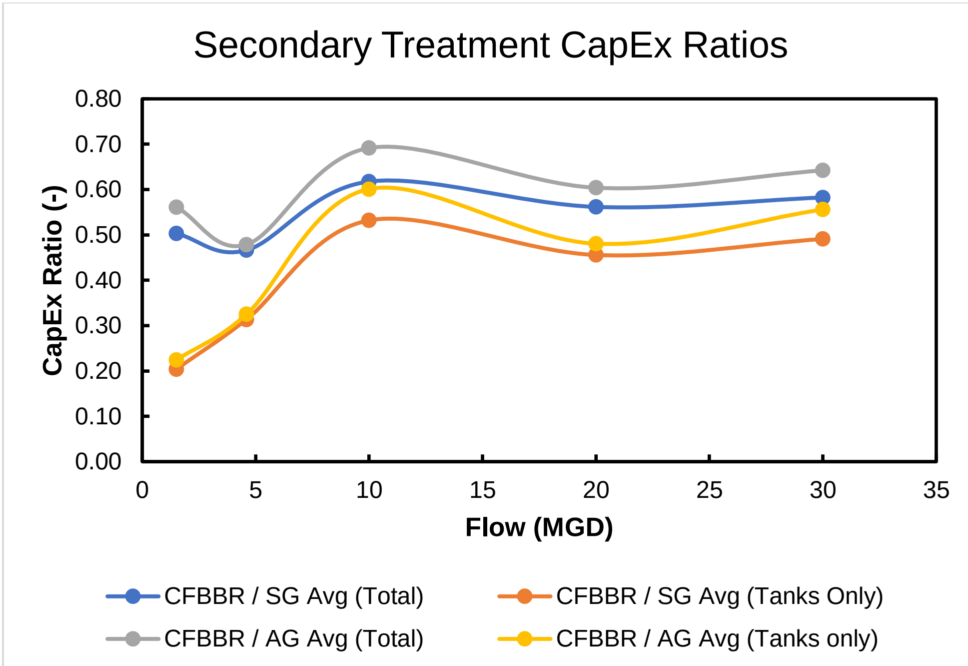

Figure 4-5b CapEx ratios for secondary treatment

The CapEx required for anaerobic digestion was far lower for the CFBBR due to lack of primary sludge production and the low observed biomass yield from the CFBBR. **Appendix 1 - Tables A1.2 – A1.6** detail the biosolids production at each treatment stage, showing that a lack of primary sludge production and reduced yield for the CFBBR reduced total biosolids from the liquid by roughly 50% compared to the other processes. As a result, AD CapEx savings were consistently around 50% for all influent flows. However, as waste activated sludge has a lower digestibility than primary sludge, especially when digested alone, and the influent COD loadings to the CFBBR are higher than other processes due to lack of primary clarification, the CFBBR's mass rate of digested biosolids is close to that of the other processes(Metcalf and Eddy, 2003). From **Appendix 1 - Tables A1.2 – A1.6**, the final digested and dewatered biosolids mass rates are fairly similar between all processes, with the CFBBR falling within the middle of the range of dewatered solids mass rates.

4.3.4 OpEx

CapdetWorks showed an advantage for the CFBBR at 1.5 and 4.6 MGD for total OpEx, however, the CFBBR became comparable to the other process at 10 MGD and above, as shown in **Figure 4-3b**. The breakdown of the OpEx components is shown in **Appendix 1 - Figures A1.6 – A1.10**. Some of the largest and most important contributors to the CFBBR's OpEx were the aeration energy, liquid pumping fluidization energy, and chemical costs for phosphorus removal, tabulated in the **Appendix 1 - Tables A1.7 – A1.11**. The fluidization energy alone contributed 50%-60% of the total energy consumption for the CFBBR plants and aeration contributed 30% of the CFBBR plants' total energy demand. Furthermore, the increase in chemical phosphorus removal for the CFBBR contributed 9% - 15% of the CFBBR's total OpEx, compared to the 5% - 8% for other process. These were offset by the decrease in OpEx for the anaerobic digesters, and no OpEx for primary clarification.

The details of annual biosolids production are presented in **Appendix 1 - Tables A1.2 – A1.6**. As apparent from **Appendix 1 - Tables A1.2 – A1.6**, the CFBBR produced 5%-15% less solids than other attached processes and comparable to the suspended growth processes. Anaerobic digestion OpEx for the CFBBR was 30%-40% lower compared to the average of the suspended and attached growth technologies. Furthermore, forgoing primary clarification reduced overall preliminary & primary (pumping, screening, grit removal, and primary clarification) OpEx by 10%-20% for the CFBBR compared to the other processes. However, due to the reduced digestibility of the biosolids (TWAS only for the CFBBR), the subsequent volume of digested biosolids was similar to the other processes and so, no significant reduction in dewatering OpEx was predicted. Despite this, the CFBBR was on the low-end in terms of total biosolids production. This an important consideration for costs and environmental footprints. With its low biosolids production, the CFBBR's contribution to landfills will be lower than the other processes, and fuel consumption by the trucking of biosolids will in turn be lower as well.

The higher aeration requirements are due to the CFBBR treating a higher COD loading, as no COD was removed by primary clarification. Furthermore, the low biomass yield means that more oxygen is consumed for COD removal as the remaining COD is converted to CO_2 instead of biomass.

The fluidization energy is unique to the CFBBR. While it can be reduced by using lighter media, this has been shown to compromise treatment capability due to the typically poorer attachment characteristics of lighter media (L. Wang et al., 2021). Across all scales, the OpEx for the fluidization energy accounted for approximately 10% of the total secondary treatment OpEx. Fortunately, the OpEx savings from primary clarification and anaerobic digestion, as well as reduced material maintenance costs offset the fluidization energy cost at lower scales.

Overall, the aeration energy accounted for 8% - 10% of the total secondary treatment OpEx for the CFBBR-based plants, similar to other processes. However, the fluidization energy contributed 12%-15% of the CFBBR's OpEx, compared to the other process where RAS and other liquid pumping contributed 6%-8% of the secondary treatment OpEx. The full breakdown of energy demand for each process is detailed in **Appendix 1 – Tables A1.7 – A1.11**.

The CFBBR has also higher demand for chemical P precipitation. This higher loading was caused by two factors: no particulate phosphorus removal via primary clarification and the CFBBR's low biomass yield. Due to the lower yield, less phosphorus was taken up by biomass growth, therefore leaving a higher amount of soluble phosphorus after biological treatment. Combined, these factors left a higher amount soluble phosphorus to be removed via precipitation and resulted in a 20%-30% increase in OpEx for the phosphorus removal compared to the non-EBPR processes, i.e., MLE, 4-BDP, MBBR, IFAS and BAF, across all flows.

4.3.5 Net Present Value

To further analyze the CFBBR's cost effectiveness the net present value was calculated for a 30-year lifespan at 5% interest, using the total CapEx and OpEx. The NPV results at 5% can be seen in the **Appendix 1 - Figures A1.11-A1.15** and in **Table 4-3**. The NPV's were also calculated at interest rates of 3% and 7% for a sensitivity analysis, also tabulated in **Table 4-3**, but there were no major differences in how the processes compared by NPV. At all interest rates, the CFBBR had the lowest NPV at 1.5 MGD and 4.6 MGD and was 3rd-5th lowest at 10 MGD to 30 MGD.

Table 4-3 Summary of averages and standard deviations of CapEx, OpEx, and NPV

Process	Cost	Flow									
		1.5		4.6		10		20		30	
		Avg	Std.Dev.	Avg	Std.Dev.	Avg	Std.Dev.	Avg	Std.Dev.	Avg	Std.Dev.
SG Avg	CapEx	7,900	620	19,300	676	36,400	1,500	64,700	4,180	94,500	6,140
	OpEx	1,950	175	4,180	445	7,210	831	12,400	1,830	17,700	2,650
	NPV 3%	46,200	4,030	101,000	9,330	178,000	17,600	308,000	39,800	442,000	57,400
	NPV 5%	37,900	3,290	83,500	7,450	147,000	14,100	256,000	32,100	367,000	46,200
	NPV 7%	32,100	2,780	71,100	6,130	126,000	11,600	219,000	26,600	315,000	38,400
AG Avg	CapEx	8,010	226	18,100	1,670	33,100	2,920	61,100	4,970	87,300	6,150
	OpEx	1,970	98	4,140	48	7,010	199	12,600	553	17,700	840
	NPV 3%	46,500	1,770	99,200	2,330	170,000	6,810	307,000	15,800	434,000	22,600
	NPV 5%	38,200	1,360	81,700	2,170	141,000	5,970	254,000	13,500	359,000	19,000
	NPV 7%	32,400	1,070	69,400	2,070	120,000	5,380	217,000	11,800	307,000	16,500

CFBBR	CapEx	5,560	10,300	24,600	41,000	61,500
	OpEx	1,590	3,520	7,400	12,500	18,500
	NPV 3%	36,700	79,300	169,500	283,500	420,500
	NPV 5%	30,000	64,400	138,000	231,000	343,000
	NPV 7%	25,300	54,000	116,500	194,500	289,000

*SG Avg; Suspended growth average, AG Avg; Attached growth average

With respect to the other conventional processes, EBPR processes (A2O, UCT and 5-BPD) have advantages in OpEx, as shown in **Appendix 1 - Figures A1.6-A1.10**, due to their significantly lower ferric requirements for chemical phosphorus precipitation. Additionally, all three attached growth processes showed similar cost estimates across all scales and were similar in OpEx when compared to the CFBBR.

From the NPV calculations, we can see that CFBBR has the biggest advantage at the lower scale of 1.5 and 4.6 MGD, with a clearly lower NPV. However, at the higher scales of 10, 20 and 30, the CFBBR's NPV is only comparable to the other processes. **Figure 4-6** plots the NPV ratios, showing that overall, the CFBBR has cost savings at 1.5 and 4.6 MGD of 15% - 25% but becomes much closer to the other processes at 10 MGD and above (< 10% savings). It is apparent both from **Table 4-3** and **Figure 4-6** that the CFBBR offers cost savings relative to both conventional suspended growth and attached growth systems, albeit declining with the increase in flows above 4.6 MGD. The results detailed in **Table 4-3** also show the considerable contribution of the OpEx to NPV, CapEx accounting for only 18%-28% of the NPV. For the CFBBR, CapEx accounts for 17%-19% of its NPV at all scales. Therefore, significant reductions in the OpEx for the CFBBR will help improve its cost effectiveness far more than CapEx reductions will. OpEx costs that can potentially be reduced are the energy consumption, chemical addition, and solids handling and disposal, which combined account for roughly 15%-30% of the OpEx. Potential avenues for OpEx reduction are discussed further in section 3.6.

Ultimately, the NPV calculations confirmed what was already discussed above about the CFBBR's potential cost effectiveness; lower CapEx across all scales, but the modularity of the system and its OpEx would hinder its cost effectiveness at higher scales. A clear potential for cost savings is demonstrated at smaller scales (< 5 MGD), but further development is required to reduce energy demand in the system to make it cost effective at larger scales ($\geq$ 10 MGD). The cost advantage at lower scales would be very important for small cities and municipalities that have nutrient discharge limits. With a CFBBR-based plant, these limits can still be met but for significantly lower costs. At larger scales, though the CFBBR is not cost competitive, it offers land footprint reduction by eliminating primary clarification and reducing the size of secondary and biosolids treatment units. Land footprint is important in site-selection and upgrades for urban treatment plants.

To further evaluate the effects of the CFBBR's modularity on the CapEx, a second costing simulation was done using a CFBBR module that can treat up to 5 MGD instead of 1.5 MGD at an HRT of 3 hours (same as the first simulations). In these simulations, the required modules for 5, 10, 20 and 30 MGD are 1, 2, 4, and 6 modules, respectively, compared to 3, 7,13, and 20. The 30-year NPV at 5% interest was recalculated and is shown in **Figure 4-6** against the NPV of the smaller module (1.5 MGD) costing simulations. **Figure 4-6** shows that at flows at 10 MGD and above, the 5-MGD module CFBBR becomes more cost competitive than the 1.5 MGD, offering on average about 17% reduction in NPV relative to the conventional suspended and attached growth process , over the entire range of flows. Thus, size limitations are one of the CFBBR's greatest hinderances to large scale (>5 MGD) plants.

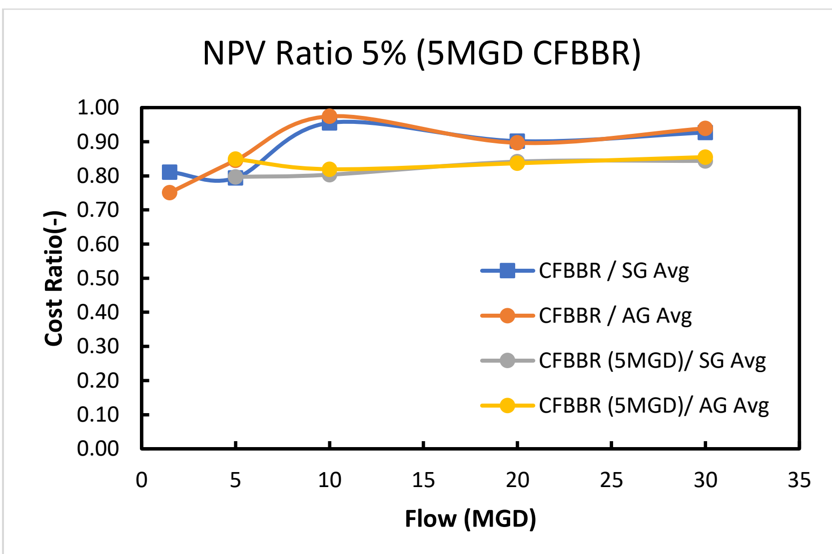

Figure 4-6 CFBBR / conventional NPV at all scales

4.4 Future Perspectives

The process modelling and costing simulations of this study highlighted several aspects of the CFBBR that leave room for improvement to further enhance its cost effectiveness, namely the fluidization energy and cost for chemical phosphorus removal. As mentioned in section 3.5 (Net Present Value), because of the OpEx's high contribution the total NPV, reduction in OpEx will be vital for lowering the CFBBR's NPV, particularly at flows above 10 MGD, to render it more cost competitive.

The CFBBR design for this study was that of an upflow fluidized bed, using particles heavier than water and fluidized upwards by liquid and air flow. There is growing interest in inverse fluidized beds for biological process such as biological wastewater treatment. In an inverse fluidized bed, buoyant particles are fluidized downwards by either downward liquid flow or upward air flow (Karamanev & Nikolov, 1992a). Studies have demonstrated that inverse fluidized bed bioreactors (IFBR) require less liquid and/or air flow to maintain fluidization compared to upright fluidized beds, greatly reducing the energy usage of the reactor(H. Wang et al., 2020). However, the surface characteristics of the buoyant particles, typically plastics, are less effective at holding biofilm compared to materials like lava rock or zeolite used in upflow CFBBR's. Surface modifications of the particles can counteract this and improve the loading capabilities of the IFBR. This was done by Wang et al. by embedding materials such as active carbon, zeolite and lava rock on the surface of polypropylene particles improving the treatment performance of the IFBR(L. Wang et al., 2021).

Enhanced biological phosphorus removal (EBPR) has been demonstrated in the CFBBR when particles are circulated between the aerobic and anoxic columns (Patel et al., 2006). Unfortunately, this could not be simulated in CapdetWorks as its attached growth process units do not have EBPR capabilities. Nonetheless, chemical phosphorus removal, which accounted for 9%-15% of the CFBBR's total OpEx and was 20%-30% higher than other processes, can be optimized. Further experimental studies and modelling in GPS-X (or other software) should be done to evaluate the potential cost savings for an EBPR-capable CFBBR. EBPR in a CFBBR would induce numerous changes throughout the plant: firstly, the Chem P OpEx would drop drastically, and the total solids produced would also decrease as far lower masses of chemical precipitants (ferric, alum, etc.) would be added to the treatment train. This reduction in solids production would reduce the biosolids processing and disposal costs as well. However, given that EBPR would necessitate carrier particle recirculation between the two columns, the aforementioned benefits may be offset by not only increased energy consumption, but also capital cost and maintenance costs associated with the high-solids handling pumps. This may offset the CFBBR's advantage in the secondary treatment CapEx costs. Modularity has already been identified as a hindrance to the CFBBR's CapEx in plants above 5 MGD. The 1.5-MGD module size is dictated by scale-up limitations of fluidized beds. One of the largest applications of liquid-solid fluidization is the LC Fining Ebullated Bed process, which has been designed with column diameters up to 4.5m(Z. M. Cheng et al., 2014). These size limitations mean that further development of fluidized bed scale-up is needed to increase the size of the anoxic column, to reduce the impact of modularity.

Two other key considerations for adopting the CFBBR is process stability and process simplicity.

Process Stability

Municipal wastewaters have fluctuations in both volume and composition. These variations occur on a daily basis and over longer time frames, largely due to seasonal water use and precipitation patterns (Metcalf and Eddy, 2003). It is well documented that excessive flows caused by wet weather events will harm the performance of treatment processes. In the case of biological treatment systems, wet weather events will lead to biomass washout, which must then be regrown to return to normal performance(Metcalf and Eddy, 2003).

Studies on the CFBBR's handling of dynamic loads demonstrated that it could maintain sufficient nutrient removal during wet weather flows. A peaking factor of 3 for 4 hours was simulated in a pilot-CFBBR, monitoring biomass levels and activity in both columns before, during, and after the simulated wet weather event. A total biomass loss of 8% - 10 % was observed in both columns and a 33% decrease in nitrification activity in the aerobic column. However, despite these losses, the effluent quality remained acceptable, with effluent $N\text{-}NH_4$ increasing from 2.0 mg/L to 3.4 mg/L and $N\text{-}NO_3$ increasing from 5.7 mg/L to 6.9 mg/L. These slight increases corresponded to the loss of biomass and activity, but within 24 hours of the wet weather event, effluent quality had returned to its previous quality, showing that the CFBBR handles wet weather events favourably and recovers from biomass washout quickly(N. Chowdhury, Zhu, et al., 2010).

Similar results were observed in a dynamic loading study on a lab-scale twin fluidized bed bioreactor, simulating a peaking factor of 4 for 3 hours. No significant change in effluent quality was observed during and after the wet weather event, further demonstrating the fluidized bed can handle wet weather events very effectively(Andalib et al., 2010b).

Process Simplicity

A major practical consideration for the adoption of a process in industry is its ease of operation. A processes operation must be as simple as possible to maintain peak performance and reduce operating costs as much as possible. More complicated processes require more highly trained operators to run them, meaning that the available workforce required will be limited. Rural areas, by their nature, have smaller workforces available, so expert operators may not be available in all cases.

In the case of the fluidized bed platform, the solids/particles are a significant operational factor to consider. Fluidization must be maintained to keep the system performing at optimal efficiency. Changes in the attached biofilm and by extension, the mass of particles, will affect the fluidization characteristics. If the particles lose biomass and the liquid velocity remains the same, the particle may be washed out. If the particles gain biomass and liquid velocity is not increased to compensate, the particles may settle and fluidization will cease(L. Wang, Zhu, et al., 2021). The CFBBR will require close monitoring and control in order to maintain continuous fluidization.

Furthermore, in the case of the circulating platform, additional solids handling is needed to transfer the particles between the columns (N. Chowdhury et al., 2009a). This will increase cost due to the needs solids handling equipment such as pumps or conveyors as well as monitoring of the transfer rate as the mass flow rate between columns will need to be optimized for biological nutrient removal.

To conclude, inverse fluidization and enhanced biological phosphorus removal offer potential cost reduction opportunities for the CFBBR, particularly for OpEx costs, but must be balanced with their inherent disadvantages to determine their true cost benefits.

4.5 Summary and Conclusions

The specific conclusions that can be drawn are as follows:

- The method of estimating costs with CapdetWorks coupled with GPS-X has been verified against published US EPA data.
- The CFBBR’s greatest advantage for CapEx savings is due to the smaller secondary/biological unit sizes, elimination of primary clarifiers, reduced sludge production, and subsequent reduction in anaerobic digester size. Total CapEx savings varied from 30%-45% in the 1.5-30 MGD range.
- The greatest hinderances for the CFBBR’s OpEx are fluidization energy, aeration energy, and chemical phosphorus removal, offsetting savings from forgoing primary clarification and reduction in anaerobic digestion OpEx. Chemical phosphorus removal OpEx for the CFBBR was 20%-30% higher than competitors. The energy demand for fluidization accounted 50%-60% of the CFBBR plants’ total energy demand.
- Despite these hinderances, energy savings from primary clarification elimination, and reduced needs for anaerobic digestion and material maintenance offset the OpEx hinderances and kept CFFBR economically competitive at all flows examined: 1.5-30 MGD.
- NPV estimates demonstrate that the CFBBR is most cost advantageous, on a percentage basis, at 1.5 and 4.6 MGD, but loses this advantage at higher flows, though still remains cost competitive.
- Though only cost competitive below 10 MGD, the CFBBR’s relatively smaller treatment units and elimination of primary clarification offers significant space reduction, enhancing the competitiveness of the CFBBR as an intensification technology for space-constrained large urban treatment plants.

Overall, the costing estimates and NPV calculations of the simulated processes and treatment plants have shown that the CFBBR can be a competitive option for small-scale municipal wastewater treatment both in terms of economics and treatment performance. The modularity of the CFBBR, which is dictated by the maximum size of circulating fluidized beds, is the main factor limiting its cost-competitiveness in large scale plants. The CFBBR market penetration would benefit from the use of surface-modified light plastic particles that would facilitate biofilm attachment and operation in an inverse fluidized bed mode, which would also facilitate carrier particle recirculation for enhanced biological phosphorous removal and subsequent reduction in chemical costs.

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Chapter 5

Expanded mineral materials as carrier media in an inverse three-phase fluidized bed bioreactor for wastewater treatment

Abstract

Many different materials have been used as carrier media for fluidized bed bioreactors and other attached growth wastewater treatment processes. These materials are key to the systems treatment performance and energy consumption. Natural materials, like lava rock and zeolite, have been shown to have better biofilm and treatment performance characteristics than plastic carriers, but often are heavy and thus consume more energy for fluidization. Lighter materials like plastics have been used in upright and inverse fluidized bed bioreactor applications, but these plastics have not shown as good of treatment performance compared to heavy natural carriers, typically only handling about 50%-75% of the loadings compared to upright fluidized bed bioreactors. One type of material that has not been tested is buoyant natural materials, such as expanded mineral pellets. To explore this, expanded clay was used a biofilm carrier in an inverse three-phase fluidized bed bioreactor treating synthetic wastewater via aerobic BOD removal and nitrification. At an OLR and NLR of 2.2 and 0.21 kgN/m^3/d, it achieved COD and ammonia removal efficiencies of 93% and 98%, respectively. These results compared well with previous inverse fluidized bed bioreactor studies using plastic carriers, however, the amount of suspended biomass accumulated in the column indicated that the system was not a purely fixed-film process but instead a hybrid attached and suspended growth.

5.1 Introduction

Fixed-film processes are used for many different wastewater treatment applications. (Metcalf and Eddy, 2003) Fixed-film processes have much higher biomass concentrations than suspended growth processes, giving them numerous advantages including handling higher organic and nutrient loadings, managing dynamic and shock loadings better than suspended growth processes, smaller reactor sizes, and lower biosolids production. Fixed-film processes like biological aerated filters and other biofilters are shown to have superior effluent suspended solids concentrations compared to suspended growth processes(Metcalf and Eddy, 2003).

The circulating fluidized bed bioreactor is a fixed-film process that uses fluidized particles as a biofilm carrier(Zhu et al., 2000). It is a two-column process with an aerobic and an anoxic column working in sequence with liquid circulation between the two. The purpose of liquid circulation is to facilitate nitrification and denitrification to achieve complete nitrogen removal(Patel et al., 2006). Particle circulation between the two columns has been proven to facilitate enhanced biological phosphorus removal, due the biofilm being transferred between the aerobic and anoxic/anaerobic zones(Andalib et al., 2010b). An anaerobic fluidized bed bioreactor platform (AnFBR) has also been tested for different wastewater treatment applications, including anaerobic digestion of biosolids and anaerobic treatment of thin stillage (Andalib et al., 2013). For these applications, a single fluidized bed column fluidized by liquid flow is used. Like the two-column aerobic/anoxic applications, the single anaerobic column applications have also demonstrated capability of handling higher loadings compared to the conventional anaerobic processes(Nelson et al., 2017).

One of the most important design choices for fluidized bed bioreactors and all other fixed-film processes is the carrier media. Commonly used materials include rock and other mineral-type materials, plastics, activated carbon, and zeolite(Metcalf and Eddy, 2003). These materials can be classified as synthetic or natural, and settling or buoyant. Settling (heavy) particles are used in BAF's, biofilters, IFAS, and upflow fluidized bed bioreactors, whereas MBBR's and inverse fluidized bed bioreactor use buoyant particles fluidized by aeration or liquid downflow.

The anaerobic fluidized bed bioreactor is one the earliest applications of liquid-solid fluidization for wastewater treatment and has been extensively explored for many different high-strength organic wastewaters. **Table 5-1** provides a list of anaerobic fluidized bed applications, the wastewater source, carrier material, and performance of the system.

Table 5-1 Summary of anaerobic treatment process applications of the fluidized bed bioreactor

Media Properties				Feed Type	ORL (kg/m^3/d)	COD rem %	TSS rem %	VSS rem %	Source
Carrier Material	Density (g/cm^3)	Diameter (microns)	Surface Area (m^2/g)						
Anaerobic Digestion of Biosolids									
HDPE	1.55	600-858	0.86	PS	18	62		63	(Z. Wang et al., 2016)
	1.55	600-859	0.86	TWAS	8.3	56		50	(Z. Wang et al., 2016)
	1.55	600-860	0.86	TWAS + ultrasonication	5.1	65		63	(M. M. I. Chowdhury et al., 2017)
Zeolite	2.36	600	26.5	PS	9.5	79		70	(Mustafa et al., 2014)
	2.36	600	26.5	TWAS	19	68		56	(Mustafa et al., 2014)

	2.36	600	26.5	Thin Stillage	29	88	78		(Andalib et al., 2013)
Anaerobic treatment/digestion of high-strength organic wastewater									
Perlite	0.2	1000	7.03	Distillery	35	84	-	-	(Sowmeyan & Swaminathan, 2008)
GAC[1]	0.56 (bulk)	600	-	Municipal	1.3 (35 °C)	74	-	-	(Gao et al., 2014)
	0.56 (bulk)	600	-		1.2 (25 °C)	67	-	-	
	0.56 (bulk)	600	-		1.4 (15 °C)	51	-	-	
Expanded clay	1.5	2800-3350		Sugarcane vinasse and cheese whey	40	17	-	-	(Ramos & Silva, 2018)
SIRAN	1.83	200-1500-	0.153	Cutting-oil	13	95	-	-	(Perez et al., 2007)
SIRAN	1.83	201-1500	0.153	Cutting-oil	51.3	67.1	-	-	
AC920	-	-	680	Brewery	4	88	-	-	(di Biase et al., 2018)
Zeolite	2.12	200-500	-	Distillery	10.5	77	-	-	(Fernández et al., 2008)
Silica sand	-	1000-1500	-	Acidic Mine & Hospital	4.35	96	-	-	(Makhathini et al., 2020)
GAC	0.56 (bulk)	600	438.9	Starch wastewater	83	92	-	-	(RANGASAMY et al., 2007)
Sand	-	300-500	-	Palm Oil Mill WW	10	78	-	-	(Borja & Banks, 1995)
GAC	-	600-2000	-	Cold-rolling	0.58	90	-	-	(H. H. Cheng et al., 2018)
SIRAN	1.83	1500-200	0.153	Wine Distillery	32	82	-	-	(Pérez et al., 1999)
Carbon felt	0.11	-	0.7	Synthetic	6.34	88	-	-	(Y. Yang et al., 2004)
Saponite	0.55	400-800	200	Sunflower Protein Isolates	9.3	80	-	-	(Borja et al., 2001)
Perlite	0.28	968	-	Wine Distillery	11.3	85	-	-	(Garcia-Calderon et al., 1998a)
Extendosphere	0.7	169	0.0355	Brewery	70	90	-	-	(Alvarado-Lassman et al., 2008)
Polyethylene	0.93	360	0.0166	Brewery	10	90	-	-	

[1]10x30 Mesh activated carbon

The properties of the different carriers varied widely, particularly the specific surface area which ranged from 0.02 m^2/g to 680 m^2/g. However, the organic loading and COD removal of the different studies are relatively similar with some indications that the surface area of the carrier correlates to performance. The studies using SIRAN sintered-glass pellets treated some of the highest COD loadings with good COD removal(Pérez et al., 1999; Perez et al., 2007), but Extendospheres also treated high loadings with good performance while having one of the lowest surface areas(Alvarado-Lassman et al., 2008). Therefore, high surface area alone is not indicative of performance as the area available for biomass growth and its surface roughness dictate performance.

A more recent application tested was anaerobic digestion of high-solids organic wastes like municipal biosolids and thin stillage(Andalib et al., 2013; M. M. I. Chowdhury et al., 2017; Mustafa et al., 2014; Z. Wang et al., 2016). The two materials used in these studies were high-density polyethylene and zeolite. The main differences between these materials are the HDPE is lighter than zeolite (1500 g/L < 2360 g/L) but zeolite has considerably higher specific surface area than HDPE. The zeolite showed higher VSS destruction for both primary sludge and TWAS compared to the HDPE. The zeolite achieved higher VSS destruction (56%) compared to the HDPE (50%), while also treating over double the OLR compar(Andalib et al., 2013; Z. Wang et al., 2016). In the case of high-solids wastewater treatment, a clear advantage for high surface area carriers is seen, since the zeolite showed better performance compared to the HDPE.

Table 5-2 details the applications of the FBBR to aerobic and anoxic wastewater treatment processes, specifically BOD removal, nitrification, and denitrification. Upright fluidized bed applications used both natural and synthetic materials, while the inverse applications used only plastics(Nelson et al., 2017; H. Wang et al., 2020; L. Wang et al., 2021). Upright FBBR's designed for complete biological nitrogen removal were operated with ThCOD loading ranges of 2.0 to 8.5 kgCOD/m^3/d, while inverse FBBR's handled ThCOD loadings of 2.4 to 3.0 kgCOD/m^3/d. It should be noted that the lowest performing upright FBBR used a polypropylene carrier(Liu et al., 2019), while the others used natural materials, like lava rock and chalk(Andalib et al., 2010b, 2010a; Green et al., 2001; Patel et al., 2006).

Table 5-2 Summary of wastewater treatment process applications of the fluidized bed bioreactor and the moving bed bioreactor

Media Properties				Maximum Loadings (kg/m^3/d)			Percent Removal (%)			Source
Material	Density (g/cm^3)	Diameter (μm)	Feed Type	COD	TN	ThCOD	COD	TN	N-NH_4	
Single Column - COD/BOD + TN Removal										
Sponge cubes[I]	-	3000	Synthetic	-	2.5[1]	11.4	-	-	70	(Tokutomi et al., 2010)
Chalk ($CaCO_3$)[U]	-	500-1000	Synthetic	-	1.44	6.6	-	-	93	(Green et al., 2001)
HDPE[U]	1.55	600-850	Synthetic	-	4.8 (2.68*)	12.4	-	-	54	(H. Wang et al., 2019)
Sand[U]	2.65	830	Synthetic	-	1.1 (1.03*)	4.7	-	-	94.2	(Aslan & Dahab, 2008)
Polypropylene	1.63	1570	Pesticide	7.4	6.8			76	89	(Ge et al., 2021)
Polypropylene[2]	0.95-0.99	7000x9000	Synthetic	1.2	0.14	-	88.2	86.4	95.7	(B. Wang et al., 2012)

Polyurethane sponge-carbon fibre comp.	-	15000	Coking Wastewater	1.1	0.06	-	83.4	45.5	85.3	(Li et al., 2019)
Two Column - COD/BOD + TN Removal										
Lava Rock[U]	2.56	600	Synthetic	3.3	0.34	4.8	90	71	97	(Patel et al., 2006)
	2.56	600	Synthetic	5.3	0.54	7.8	91	59	95	(N. Chowdhury et al., 2008)
	2.56	600	Synthetic	2.3	0.25	3.4	96	82	98	(Andalib et al., 2010a)
	2.56	600	Municipal	2.26	0.26	3.4	96	82	97	(Eldyasti et al., 2010)
	2.56	600	Rendering	7.3	0.27	8.5	96	96	98	(M. N. Islam et al., 2012)
Polypropylene – AC coating[I]	0.904	3200	Synthetic	2.1	0.21	3.1	84	75	99	(H. Wang et al., 2020)
Polyethylene – zeolite coating[I]	0.946	3500	Synthetic	1.64	0.17	2.4	91	40	93	(L. Wang et al., 2021)
Polypropylene[U]	1.328	1390	Septic Effluent	1.29	0.15	2.0	96	82	93	(Liu et al., 2019)

[1]N-NH_4 removed (kg/m^3/d); [2]Cylinderical particles

[U]Upflow fludized bed bioreactor; [I]Inverse fluidized bed bioreactor

Another fixed-film process that utilizes a moving carrier media is the moving bed bioreactor (MBBR). The MBBR has been used far more extensively on large scale applications than the CFBBR. Applications of the moving bed bioreactor reactor including municipal and industrial wastewater, are summarized in Table 3. Unlike the FBBR which uses both buoyant and settling carriers, MBBR’s use buoyant plastic media as carriers but are non-granular or irregular in shape (Figure 1) and generally much larger compared to fluidized bed carriers. Some applications have utilized plastic materials made from polyurethane(Chu et al., 2014; Nhut et al., 2020). MBBR’s have demonstrated comparable performance to suspended growth processes like activated sludge but as a fixed-film produce have also been shown to produce less waste biosolids(Saidulu et al., 2021) Like their suspended growth counterparts, MBBR’s are capable of nitrification and denitrification. In the case of anoxic/anaerobic MBBR’s the media is mixed by mechanical mixing instead of aeration (Metcalf and Eddy, 2003). The Norwegian company Hias patented a 3-stage MBBR process that achieves complete nitrogen and phosphorous removal by mechanically conveying carrier particles between anaerobic, anoxic and aerobic tanks to achieve nitrification, denitrification and enhanced biological phosphorus removal (Sorenson et al., 2019).

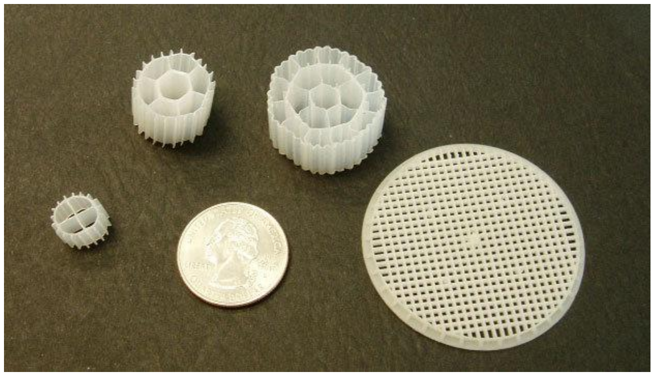

Figure 5-1 Examples of Kaldnes carrier media

Table 5-3 Summary of MBRR operating parameters and performance

Feed type	Design and Operation					Removal Efficiency		Source
	Carrier Material	Fill Ratio (%)	HRT (h)	OLR ($kg/m^3/d$)	NLR ($kg/m^3/d$)	COD	TN	
Municipal WW	Polyethylene	50	6	1.3	0.12	77.1	89.9	(X. J. Wang et al., 2006)
Laundry WW	Kaldnes-5	-	24	4.7	-	89-94	-	(Bering et al., 2018)
Domestic WW	Polyurethane sponge	20	3.6-9	0.6-1.2	0.15-0.3	72.1-97.9	67.2-70.6	(Nhut et al., 2020)
Dairy WW	Kaldnes-1	20-40	8	6.25	-	95	-	(Santos et al., 2020)
Greywater	Biomedia PZE	14	4	2.9	-	70.	-	(Chrispim & Nolasco, 2017)
Poultry WW	Polyethylene granules	-	9	-	-	94.1	50.8	(Baddour et al., 2016)
Municipal WW	Polyurethane foam	25	-	-	-	80-86	25-48	(Chu et al., 2014)
Municipal WW	Kaldnes-3	50	12	1.2	0.1	86.3	52.3	(Zinatizadeh & Ghaytooli, 2015)
Municipal WW	FLOCOR-RMP	50-70	20	-	-	96.9	84.6	(Kermani et al., 2008)
Municipal WW	Kaldnes-1	60	3-5	-	-	78-98.4	-	(Ødegaard, 2006)
Municipal WW	Kaldnes-3	33	13	0.3-0.5	-	84-96	96-99**	(Kora et al., 2020)
Hospital WW	Kaldnes-1	30	24	0.85	0.04	95.6	-	(Shokoohi et al., 2017)

*N-NH_4 removal

From the applications shown in **Tables 5-1 and 5-2**, heavy (settling) natural materials and heavy and light (buoyant) plastics have been used for fluidized bioreactor reactor applications. However, buoyant natural materials have not been tested. Some expanded materials such as perlite and clay have been tested but the clay particles were(Garcia-Calderon et al., 1998a; Ramos & Silva, 2018; Sowmeyan & Swaminathan, 2008) heavier than water. The literature does indicate that natural materials have advantages in biofilm growth characteristics over plastics, however, plastics, being lighter, require less liquid circulation for fluidization, and reduc(H. Wang et al., 2020). So, a light-weight natural material could potentially have both of the two aforementioned advantages of high biomass retention and low fluidization energy.

The purpose of this work is to evaluate several different expanded mineral-type materials as carrier media for inverse aerobic fluidized bed bioreactors. The specific objectives are : a- determine the optimal material that can sustain inverse fluidization; b- evaluate COD removal and nitrification performance; and c- benchmark performance against upright and inverse fluidized bed wastewater studies. Materials and Methods

5.2 Materials and Methods

5.2.1 Bioreactor Design

To explore the use of light natural materials as carrier media, several materials were selected for use in four parallel inverse fluidized bed bioreactors (IFBR) set in cylindrical plexiglass columns. Each column functioned independently for the purpose of conducting simultaneous tests of different reactor configurations. Each column was 7 cm in diameter (I.D.) and 100 cm tall. The liquid level was maintained at 85 cm for a working volume of 3.2 L. Each column had a stone fine-bubble aerator (LX-760, Ylong, China) installed at the base of the column and was connected to independent air hoses. Aeration was controlled by manual valves (Kitz No. 58, Japan) and measured using air flow meters (LZB-3WB, Chengfeng Flowmeter, China). Influent was fed to the top of the column by peristaltic pumps (Masterflex No. 7553-70, Antylia Scientific, USA) and effluent left via a port at the base of the column. Liquid flow was controlled by regular calibration of the peristaltic pumps. A steel wire screen (square grid, 1.5 mm x 1.5 mm) was placed 5 cm above each aerator to prevent particles from escaping or settling at the base and protect the aerator from plugging.

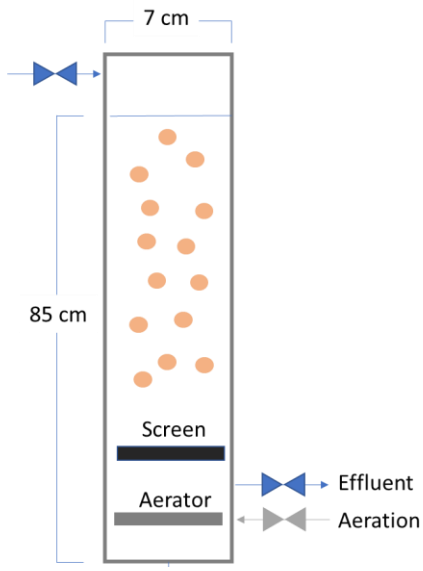

Figure 5-2 Diagram of inverse fluidized bed bioreactor

5.2.2 Bioreactor Start-up

Three materials were selected for initial testing: expanded clay (GrowIt), expanded glass (Poraver), and perlite. Three columns (R1-3) were filled to 20% of the working volume (85 L) with each material and seeded with activated sludge from the Adelaide Pollution Control Plant in London, Canada. The columns were run with no influent (aeration only) for 5 days to allow the microbes to attach to the particles without being washed out. Influent feeding of the synthetic wastewater then began at 10 L/d.

Table 5-4 Operating parameters during start-up phase (14 days)

Parameter	Unit	R1	R2	R3
Carrier		Perlite	Poraver (glass)	Clay
Influent Flow Rate	L/d	10	10	10
HRT	h	8	8	8
Air flow	mL/min	200	350	200
DO	mg/L	2.3 ± 0.5	6.7 ± 1.5	2.9 ± 1.5
Coarse bubbling		No	No	No
Bulk fill	%	20	20	20
Bed mass	g	95	150	170
EBCT	h	1.54	1.54	1.54
OLR	kgCOD/m^{3}d	1.1	1.1	1.1
NLR	kgN/m^{3}d	0.2	0.2	0.2

Volumetric ThCOD Loading	$kgCOD/m^3d$	2.1	2.1	2.1
ThCOD loading by media surface area	$gCOD/m^2d$	4.6	4.6	4.6

5.2.3 Bioreactor Operation

After the initial start-up period, the expanded clay was selected for further testing. Expanded clay was selected over perlite and Poraver for both performance and operational considerations. During operation, particularly with perlite and Poraver (expanded glass), particle settling occurred due to biofilm growth and water absorption. This created an issue of particles becoming too heavy to maintain fluidization, settling, and plugging the effluent port. This also occurred with expanded clay but to a far lesser extent, so it was manageable, and operation could continue.

It should also be noted that one column was run with 50% bed fill with expanded clay, but due to the internal friction of the bed, fluidization could not be achieved. To this end, the expanded clay should be tested in larger diameter columns, so as to reduce wall effects on the bed and improve fluidization characteristics.

Four columns (R1-4) with different amounts of expanded clay particles were seeded with 1L of return activated sludge (TSS = 9200 mg/L and VSS = 7200 mg/L) from the Adelaide Plant in London. After seeding , synthetic wastewater was fed for the start-up period and two phases of operation. The operating conditions for Phases 1 and 2 are shown in **Tables 5-5 and 5-6**. Three columns operated with fine bubble aeration only (R1, R3, and R4) and the fourth (R2) operated with a combination of fine bubble and coarse bubble aeration. The coarse bubble aeration was supplied by a split air stream with one line going to the fine bubble diffuser and the other going to a port at the center point of the column' base. Coarse bubble aeration was used to compare fluidization under different aeration configurations.

Table 5-5 Operating parameters during Phase 1 (69 days)

Parameter	Units	R1	R2	R3	R4
Influent Flow Rate	L/d	10	10	10	10
HRT	h	8	8	8	8
Air flow	mL/min	400	800	800	400
DO	mg/L	3.9 ± 0.7	3.4 ± 0.5	5.9 ± 0.8	2.0 ± 0.3
Coarse bubbling		No	Yes	No	No

Bulk fill	%	15	30	30	30
Bed mass	g	125	255	255	255
EBCT	h	1.15	2.3	2.3	2.3
OLR	$kgCOD/m^3d$	1.06	1.06	1.06	1.06
NLR	kgN/m^3d	0.12	0.12	0.12	0.12
Volumetric ThCOD loading	$kgCOD/m^3d$	1.6	1.6	1.6	1.6
ThCOD loading by surface area	$kgCOD/m^2d$	5.3	2.6	2.6	2.6

Table 5-6 Operating parameters during Phase 2 (77 days)

Parameter	Units	R1	R2	R3	R4
Influent Flow Rate	L/d	20	20	20	20
HRT	h	4	4	4	4
Air flow	mL/min	600	1500	1000	600
DO	mg/L	1.6 ± 0.8	5.6 ± 0.6	3.1 ± 1.1	2.0 ± 0.6
Coarse bubbling		No	Yes	No	No
Bulk fill	%	15	30	30	30
Bed mass	g	125	255	255	255
EBCT	h	0.6	1.2	1.2	1.2
OLR	$kgCOD/m^3d$	2.2	2.2	2.2	2.2
NLR	kgN/m^3d	0.2	0.2	0.2	0.2
Volumetric ThCOD Loading	$kgCOD/m^3d$	3.3	3.3	3.3	3.3
ThCOD Loading by surface area	$kgCOD/m^2d$	20	5	5	5

All tests were conducted using synthetic wastewater composed of sodium acetate, ammonium chloride, potassium phosphate monobasic, sodium bicarbonate and trace minerals combined with tap water. The synthetic influent was prepared in a 200 L barrel filled to 155 L and adding 70 g sodium acetate, 20.5 g ammonium chloride, 6 g potassium phosphate monobasic, and 50 g sodium bicarbonate. The target influent COD, N-NH4, P-PO4, and alkalinity characteristics were 350 mg/L, 35 mg/L, 12 mg/L and 450 $mgCaCO_3/L$ for, respectively. The resulting alkalinity concentration showed some variation due to the alkalinity already present in the tap water.

Sludge was not wasted separately from the base, instead accumulated biomass at the base was periodically (every 2-3 days) scoured with increased aeration (increased to 2 L/min) for 3 minutes prior to discharge via the effluent port. To quantify the excess solids leaving the system, the TSS and VSS of effluent samples taken during the scouring periods were measured.

5.2.4 Sampling and Measurement Methods

Effluent samples were collected from the tubes connected to the effluent port. Samples of the bulk liquid and bioparticles were collected using a sampling rod through the top of the open column. To measure TCOD, SCOD, N-NH_4, N-NO_3, N-NO_2, and P-PO_4, HACH Test Vial kit methods were used. To measure SCOD, samples were first filtered through membrane discs (Versapor, 0.45 µm x 47mm, Pall Laboratory) and the filtrate COD was measured using the HACH COD Test Vial. Alkalinity was measured by the titration method using a 0.02 N sulfuric acid solution (Hach Method 8221). The USEPA Gravimetric Method (standard method 2540) was used to measure TSS and VSS concentrations in the effluent and bulk liquid. To measure attached VSS on the carrier particles, a sample of particles was decanted, immersed in DI water, and then sonicated in a sonicator bath (VWR AquaSonic Model 75HT) for 3 hours to detach all solids from the particles. The VSS in the DI water was then measured by the USEPA Gravimetric Method. The mass of VSS detached from the mass of particles was then calculated and divided by the mass of dried carrier particles to yield a mass of VSS to mass of marrier value.

Specific nitrification rate (SNR) tests were performed on the carrier particles. To perform these tests, a stock solution of ammonium chloride (NH_4Cl) and sodium bicarbonate ($NaHCO_3$) was prepared with target concentrations of 50 mgN/L as NH_4-N and 400 mg/L as $CaCO_3$ for alkalinity. 1 L Erlenmeyer flasks were filled with the solution and a sample of carrier particles (approx. 15 g) were added to the solution. A fine bubble diffuser was placed in each flask to provide aeration and mixing. Samples were collected at specific time intervals (T = 0, 0.5, 1, 2, 3, 4, 5, & 6 hours) to measure NH_4-N, NO_3-N, and alkalinity using the previously specified methods. Dissolved oxygen was also measured at the same time intervals to ensure adequate aeration was provided.

Due to Covid-19 restrictions on access to laboratory space and equipment, only TCOD, NH_4-N, NO_3-N and NO_2-N could be measured during Phase 1. In the case of SNR tests performed during Phase 1, the ammonia removal rate could only be measured and normalized to the mass of media, instead of the mass of attached or suspended VSS, as TSS and VSS measurements could not be performed. For this same reason, no SNR tests were performed during Phase 2.

5.3 Results and Discussion

5.3.1 Start-Up Testing

During start-up, both perlite and Poraver presented several operational challenges. Both were washed out very frequently making it difficult to maintain the bed size and they both absorbed water to the point where they lost buoyancy and began settling in the column. This led to plugging issues and difficulty maintaining fluidization.

For these reasons, only the expanded clay continued to be used for this study as it did not present any major operational difficulties.

Beyond the operational considerations, the expanded clay showed the best nitrogen removal during the first 14 days of operation, shown in **Table 5-7**. It can be seen in **Table 5-7** that the perlite and Poraver reactors (R1 and R2, respectively) did not achieve full nitrification and showed poor COD removal and high effluent suspended solids. While the clay had not yet achieved full nitrification either, it can be seen in the concentration trends in **Figure 5-3** that the ammonia removal was continuing to improve compared to the perlite and Poraver columns. A paired t-test of the average ammonia effluents values confirmed the difference between the averages of R1 and R3 (38 mg/L - 35 mg/L) with a confidence of 90% ($\alpha = 0.1$, $p = 0.07$). It was also observed that the effluent TSS of R1 was about double that of R3 (146 mg/L vs 79 mg/L). Given these results and a multitude of plugging and particle washout issues with the perlite and Poraver columns, the expanded clay was selected for further testing.

Table 5-7 Average influent and effluent quality during start-up

Parameter	14 Days	Inf	R1	R2	R3
TCOD	mg/L	349 ± 50 (6)	125 ± 39 (6)	104 ± 18 (6)	143 ± 56 (6)
SCOD	mg/L	-	81 ± 20 (6)	37 ± 14 (6)	56 ± 17 (6)
$N\text{-}NH_4$	mg/L	71 ± 3 (6)	35 ± 9 (6)	52 ± 7 (6)	38 ± 18 (6)
$N\text{-}NO_3$	mg/L	-	2 ± 1 (6)	1 ± 0.5 (6)	12 ± 7 (6)
Alk as CaCO3	mg/L	455 ± 28 (6)	383 ± 60 (6)	487 ± 71 (6)	387 ± 138 (6)
TSS	mg/L	-	146 ±149 (4)	66 ± 16 (4)	79 ± 35 (4)
VSS	mg/L	-	95 ± 64 (4)	59 ± 19 (4)	72 ± 29 (4)

- Avg ± Std. Dev. (n-samples)

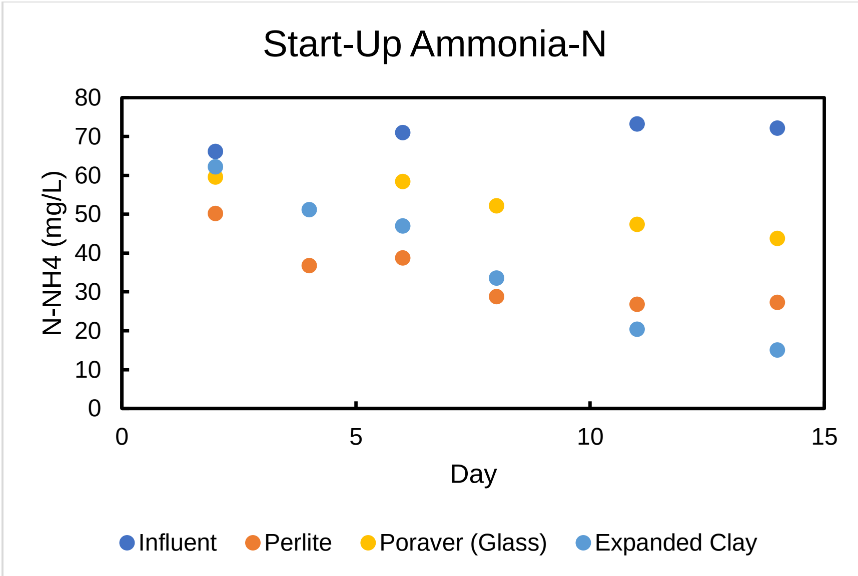

Figure 5-3 Concentration trend of ammonia during start-up testing

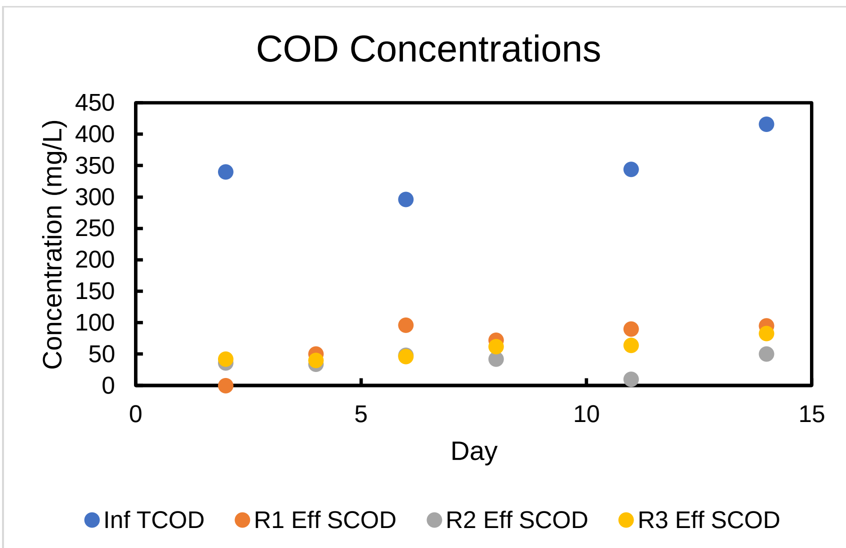

Figure 5-4 COD Removal Performance During Start-Up

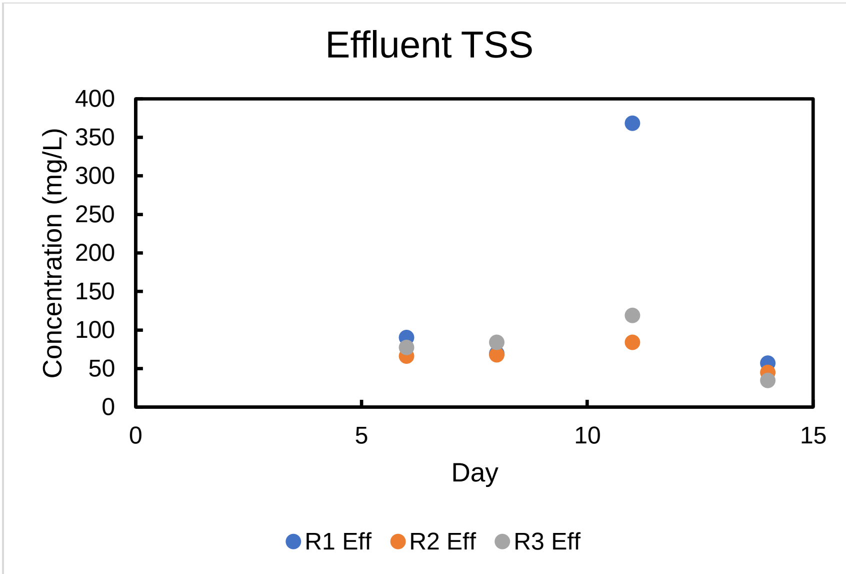

Figure 5-5 Effluent TSS during Start-Up

5.3.2 COD Removal

During Phase 1, all four columns achieved full COD removal with efficiencies above 93% and effluent TCOD concentrations below 24 mg/L. Though TSS/VSS could not be measured during this time, these low TCOD concentrations indicate that very little suspended solids were present in the effluent and bulk liquid, meaning that no downstream solids removal would be required at the Phase 1 COD loading of 1.1 kgCOD/m^{3}d.

During Phase 2, the COD removal performance was more varied than in Phase 1. Due to the high effluent TSS/VSS concentrations (>50 mg/L), the effluent TCOD concentrations were higher than desirable (70 mg/L to 160 mg/L). Downstream solids separation would be required to reduce the effluent TCOD and improve overall COD and TSS removal. R1, R2, and R3 achieved average effluent SCOD concentrations of 30 mg/L or lower, whereas R4 had an effluent SCOD of 45 mg/L. For R1, R2, and R3, the effluent soluble COD of 30 mg/L coupled with downstream solids removal would correspond to a COD removal of at least 92%.

5.3.3 Nitrogen Removal

During Phase 1, at an OLR and NLR of 1.1 kgCOD/m^3/d and 0.1 kgN/m^3/d, respectively, ammonia was completely removed via nitrification. The results in **Table 5-8** show that all four columns achieved full nitrification at steady-state , though it is observed in **Figure 5-6** that R2 and R3 reached steady earlier in Phase 1. R2 and R3's effluent ammonia stabilized after 20 days, while R1 and R4 stabilized after 48 days. As R2 and R3 had double the bed fill than R1 and R4, this would indicate that the R1 and R4 were limited by their smaller bed size and took longer to grow enough biofilm necessary to achieve full nitrification at steady-state.

Table 5-8 Average influent and effluent quality during Phase 1

Phase 1	69 Days	Inf	R1	R2	R3	R4
TCOD	mg/L	337 ± 5 (18)	6.5 ± 0.7 (18)	24.0 ± 17 (18)	14.0 ± 1 (18)	21.5 ± 18 (18)
$N\text{-}NH_4$	mg/L	36.8 ± 4 (18)	0.4 ± 0.3 (18)	0.7 ± 0.7 (18)	0.2 ± 0.2 (18)	0.7 ± 0.7 (18)
$N\text{-}NO_3$	mg/L	-	19.4 ± 2 (18)	20.3 ± 2 (18)	19.6 ± 2 (18)	17.3 ± 1 (18)
$N\text{-}NO_2$	mg/L	-	0.7 ± 0.6 (18)	0.4 ± 0.3 (18)	0.7 ± 0.7 (18)	1.2 ± 0.8 (18)

- Avg ± Std. Dev. (n-samples)

This was confirmed in Phase 2 at double the organic and nitrogen loadings, since after 77 days R1 and R4 never reached full nitrification, though R2 and R3 did. It is observed from **Figure 5-6** that R2 and R3's ammonia effluent stabilized after approximately 25 days, similar to the time it took to stabilize in Phase 1, while R1 and R4 remained unstable throughout Phase 2. Both R1 and R4 showed effluent $N\text{-}NH_4$ above 10 mg/L, $N\text{-}NO_2$ above 2 mg/L and $N\text{-}NO_3$ below 10 mg/L. R2 and R3's steady-state effluent quality was similar to Phase 1, with effluent $N\text{-}NH_4$, $N\text{-}NO_3$, $N\text{-}NO_2$ of approximately 1, 20 and 1mg/L, respectively.

Table 5-9 Average influent and effluent quality during Phase 2

Phase 2	77 days	Inf	R1	R2	R3	R4
TCOD	Avg	358 ± 32 (19)	158 ± 120 (19)	89 ± 73 (19)	73 ± 63 (19)	112 ± 82 (19)
SCOD	Avg	-	30 ± 9 (19)	25 ± 11 (19)	26 ± 12 (19)	43 ± 41 (19)
$N\text{-}NH_4$	Avg	36 ± 4 (19)	11 ± 8 (19)	0.6 ± 0.9 (19)	1 ± 1 (19)	14 ± 10 (19)
$N\text{-}NO_3$	Avg	-	6 ± 7 (19)	22 ± 3 (19)	19 ± 4 (19)	6 ± 6 (19)
$N\text{-}NO_2$	Avg	-	4 ± 3 (19)	2 ± 2 (19)	0.8 ± 0.5 (19)	2 ± 2 (19)
$P\text{-}PO_4$	Avg	13.4 ± 0.6	13.2 ± 2 (19)	12.8 ± 0.5 (19)	12.7 ± 1 (19)	12.5 ± 3 (19)
TSS	Avg	-	109 ± 101 (19)	80 ± 56 (19)	93 ± 114 (19)	98 ± 65 (19)
VSS	Avg	-	88 ± 90 (19)	58 ± 48 (19)	53 ± 51 (19)	77 ± 54 (19)
Alk as CaCO3	Avg	444.4	413 ± 51 (19)	337 ± 11(19)	335 ± 29 (19)	409 ± 52 (19)
Obs. Yield	gVSS/gCOD		0.4	0.18	0.22	0.23
Att. VSS	mgVSS/gmedia		4.2	7.9	6.9	4.8
Sus. VSS	mg/L		896.7	1619.8	886.0	724.5
SRT	d		1.5	5.2	5.6	1.5

- Avg ± Std. Dev. (n-samples)

The long-term stability of R2 and R3 is observed in **Figures 5-7 and 5-8**, given that once steady-state nitrification is achieved the effluent ammonia and nitrates show no major variance throughout the steady-state period of each phase. It is also shown in **Tables 5-8 and 5-9** that at steady-state nitrite accumulation was very low for both R2 and R3, indicating that the biofilm remained stable and did not suffer major washouts or loss of activity. **Figures 5-9 and 5-10** graph the daily ammonia removal rate against the effluent VSS concentrations for R1 and R4, respectively, with both showing a generally negative trend, indicating a loss of ammonia removal/nitrification activity due loss of VSS/biomass.

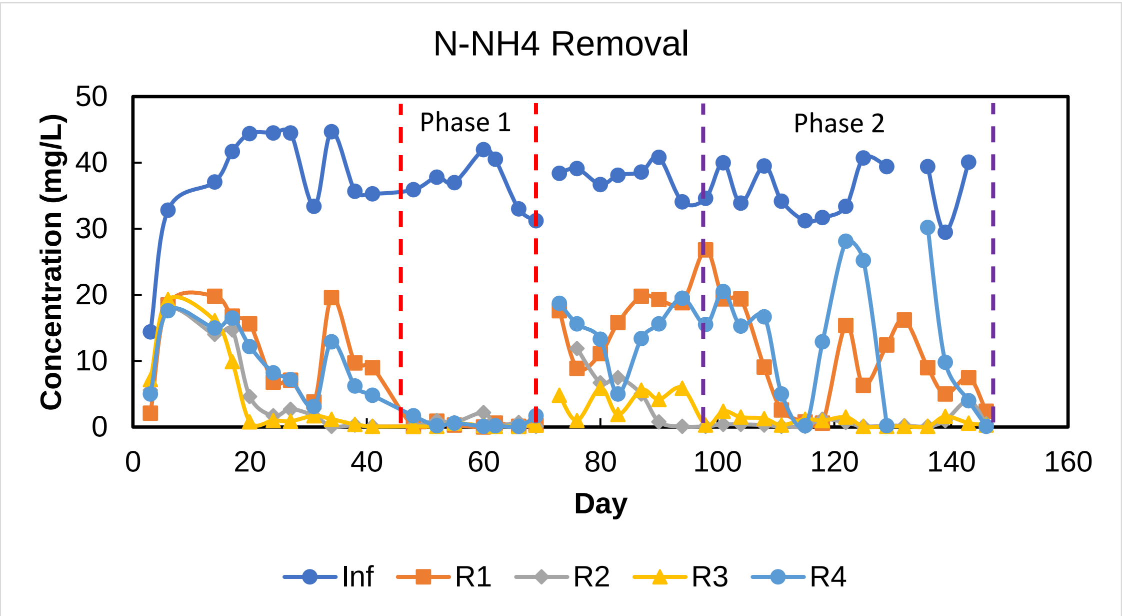

Figure 5-6 Influent and effluent ammonia concentrations

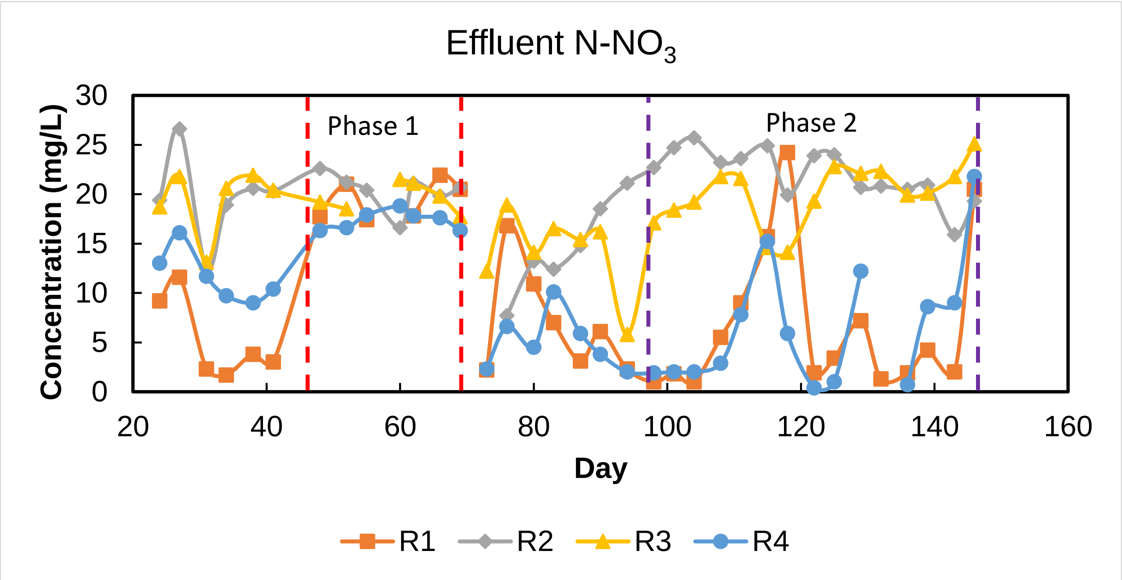

Figure 5-7 Effluent nitrate concentrations

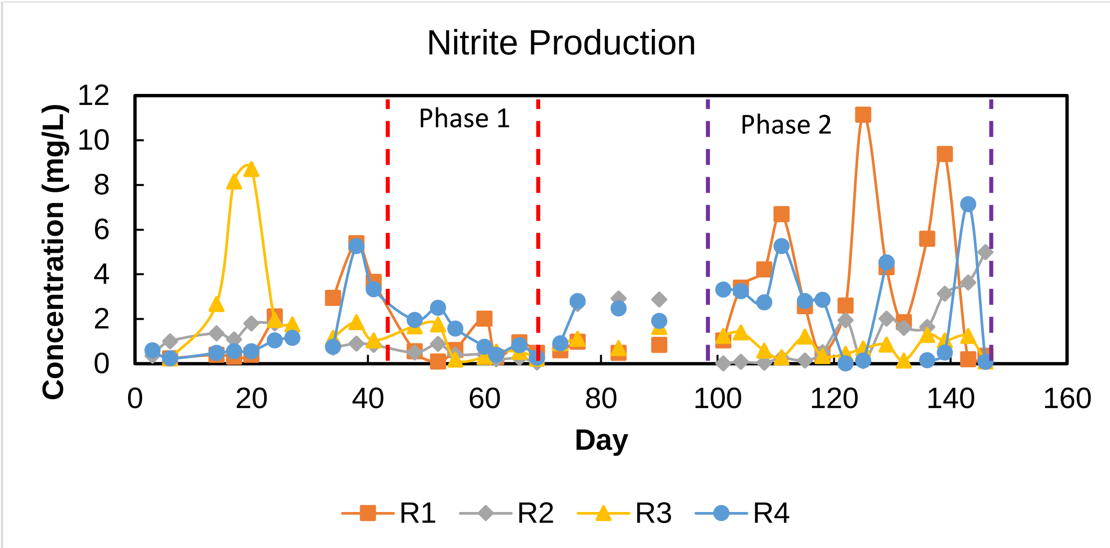

Figure 5-8 Effluent Nitrite concentrations

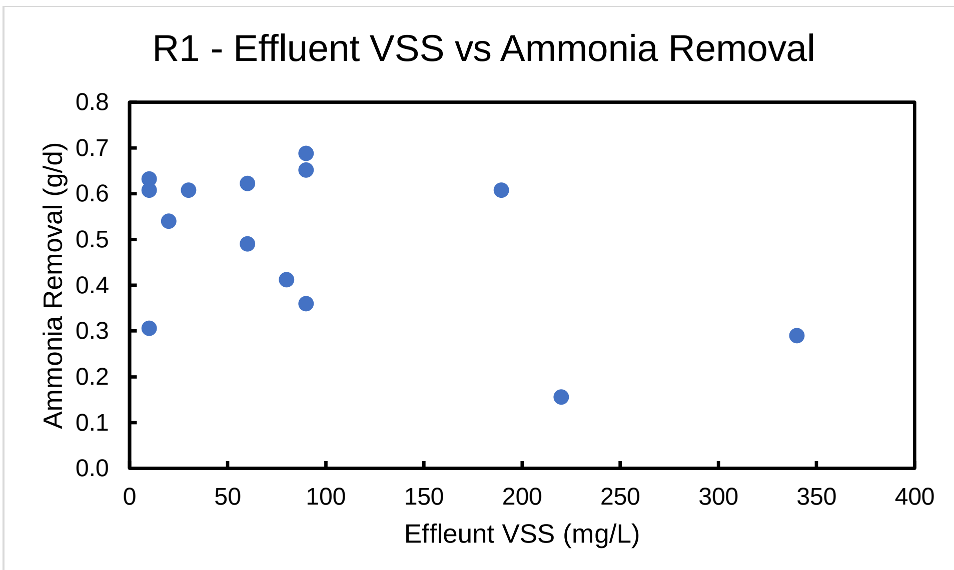

Figure 5-9 R1 Ammonia removal against effluent VSS

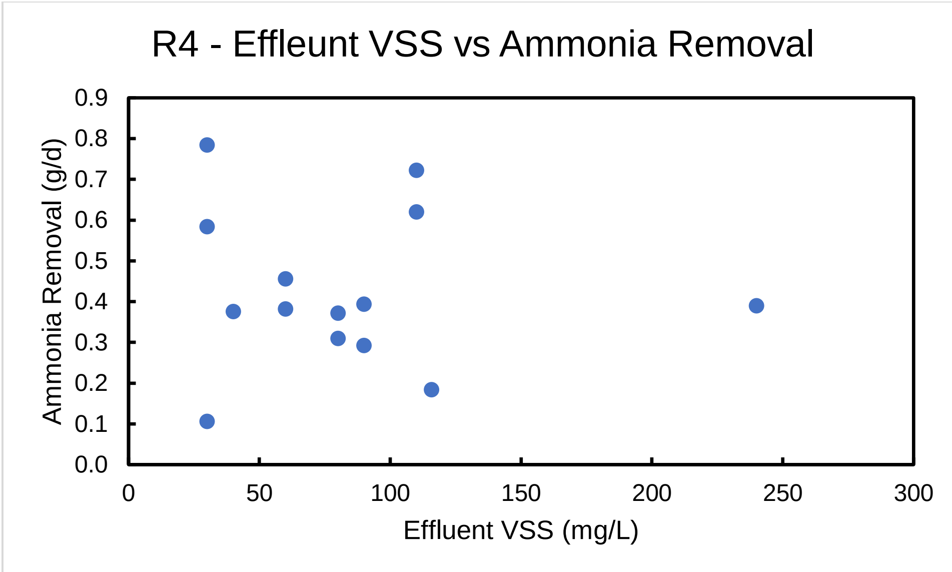

Figure 5-10 R4 Ammonia removal against effluent VSS

5.3.4 Nitrate, Nitrite, and Nitrite Accumulation Ratio (NAR) Variations

At steady-state in Phase 1, the effluent nitrogen quality was stable for all four columns. In Phase 2, R2 and R3 had relatively stable N-NH_4 removal and effluent nitrate and nitrite concentrations. R1 and R4 did not reach steady-state and showed significant variation in effluent ammonia, nitrates and nitrites, shown in **Figures 5-11 and 5-12**, respectively. The nitrite accumulation ratio, Equation 5-1, for each reactor is shown in **Figure 5-13**. At steady-state, R2 and R3 showed NAR's below 0.1, whereas R1 and R4 had NAR's mostly ranging from 0.3-0.8, further demonstrating the instability of nitrification in R1 and R4.

$$NAR = \frac{NO_2}{NO_2 + NO_3} \quad (5\text{-}1)$$

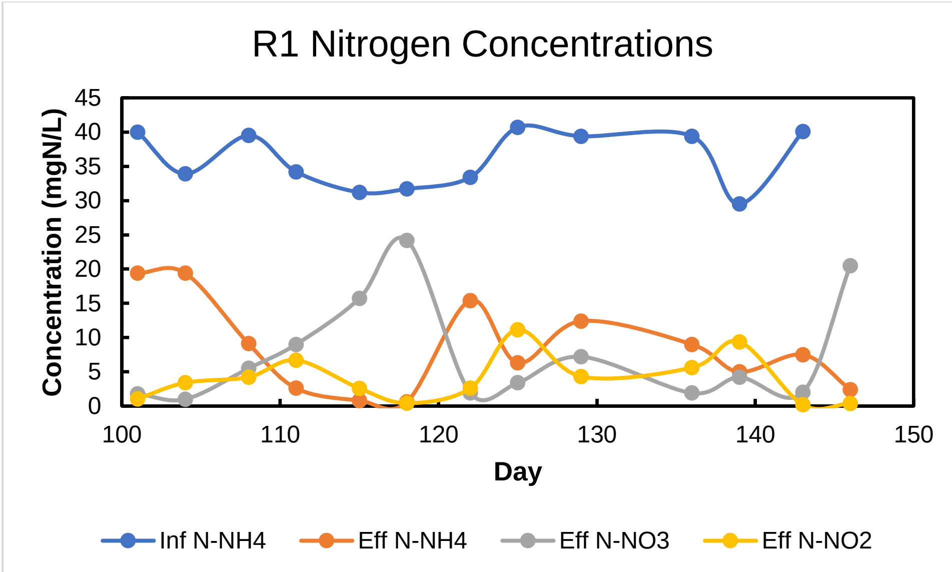

Figure 5-11 R1 Nitrogen Concentration over time

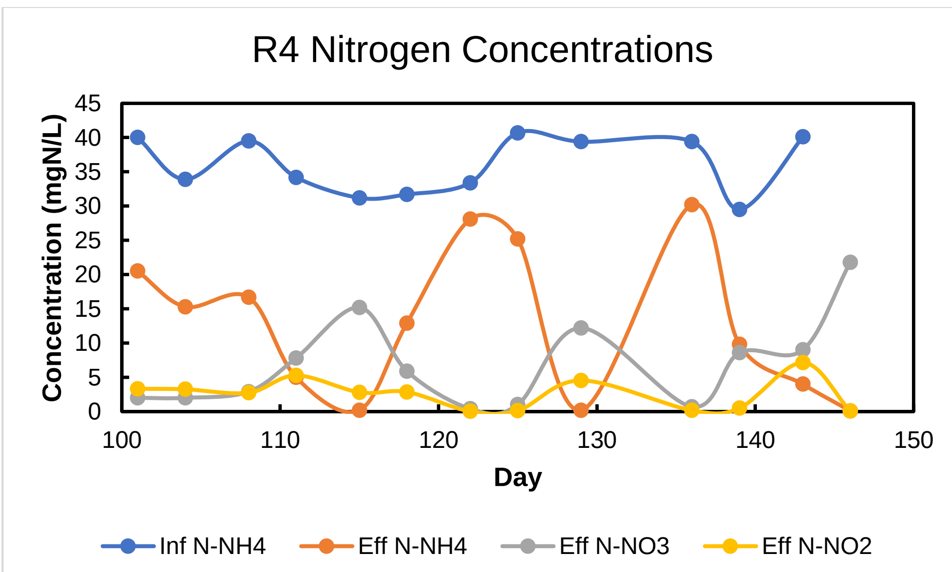

Figure 5-12 R4 Nitrogen Concentration over time

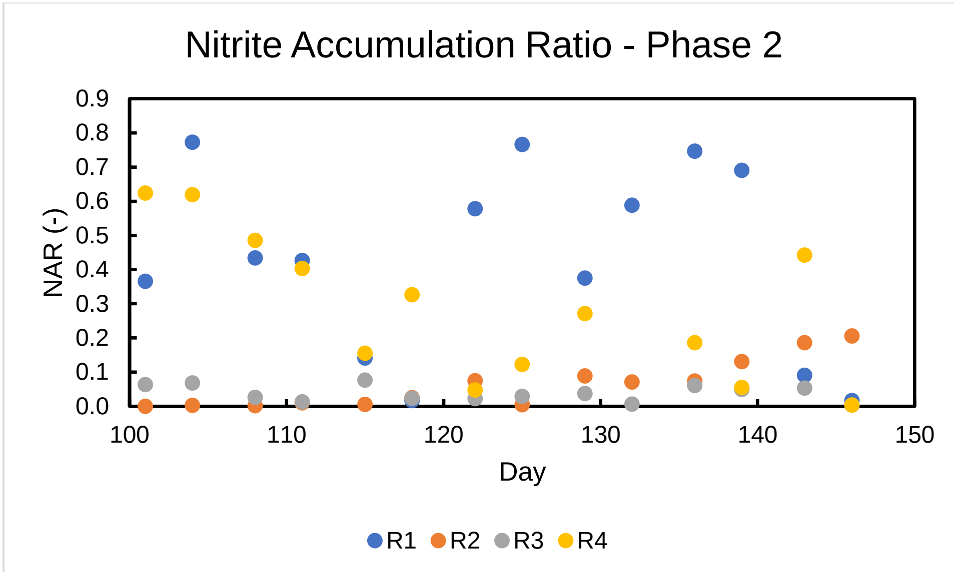

Figure 5-13 Nitrite Accumulation Ratio During Phase 2

In both figures, several increases in effluent ammonia are observed, which corresponded to drops in effluent nitrate and nitrite, indicating a significant loss in nitrification activity. The average N-NO_2/NO_x ratios for R1 and R4 were 0.4 and 0.25, respectively, which further indicates that the nitrification activity, particularly the nitrite oxidizing activity is far lower than the ammonia oxidizing activity. As a R1 and R4 both had average DO concentrations between 1-2 mg/L and smaller media masses, the nitrite accumulation is likely due to a combination of the two factors.

5.3.5 Bulk Liquid and Effluent Suspended Solids

Very high concentrations of suspended solids were observed during Phase 2. The bulk liquid suspended solids ranged from 800 mg/L to 1600 mgVSS/L and the effluent suspended solids ranged from 80 to 100 mgTSS/L. The high concentrations of suspended biomass (as VSS) indicated that the systems were not functioning as fixed-film processes, but rather hybrid attached/suspended growth processes. This is confirmed by the calculated values of total VSS, suspended and attached, for all columns. In all cases, the suspended biomass fraction represented the larger fraction of biomass, shown in **Figures 5-14 to 5-17**.

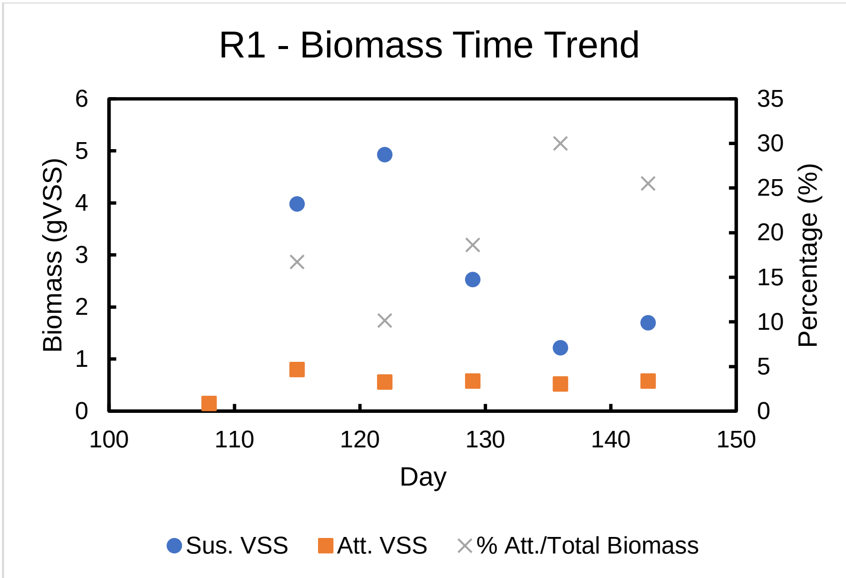

Figure 5-14 Reactor 1 biomass time trend

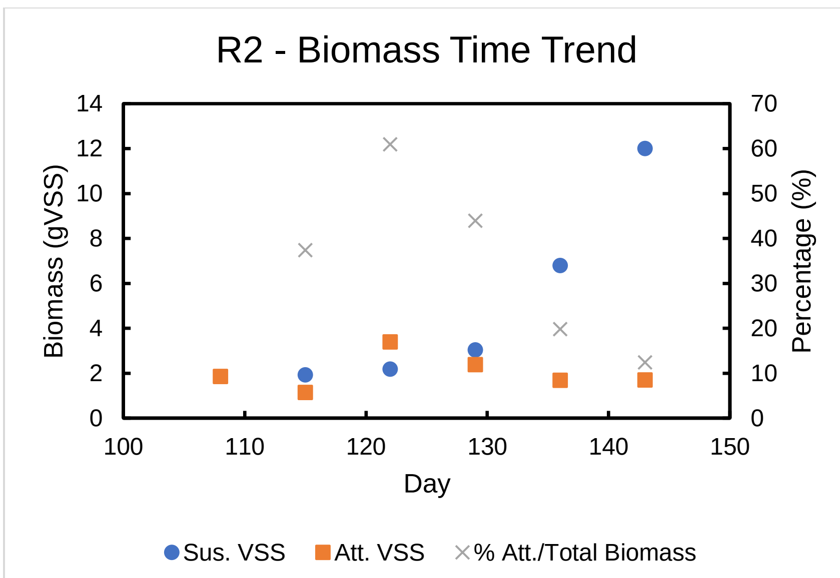

Figure 5-15 Reactor 2 biomass time trend

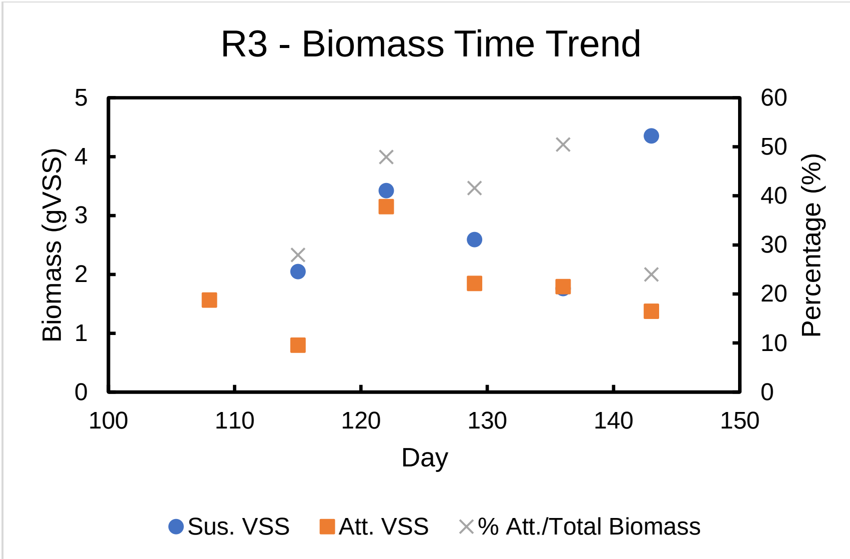

Figure 5-16 Reactor 3 biomass time trend

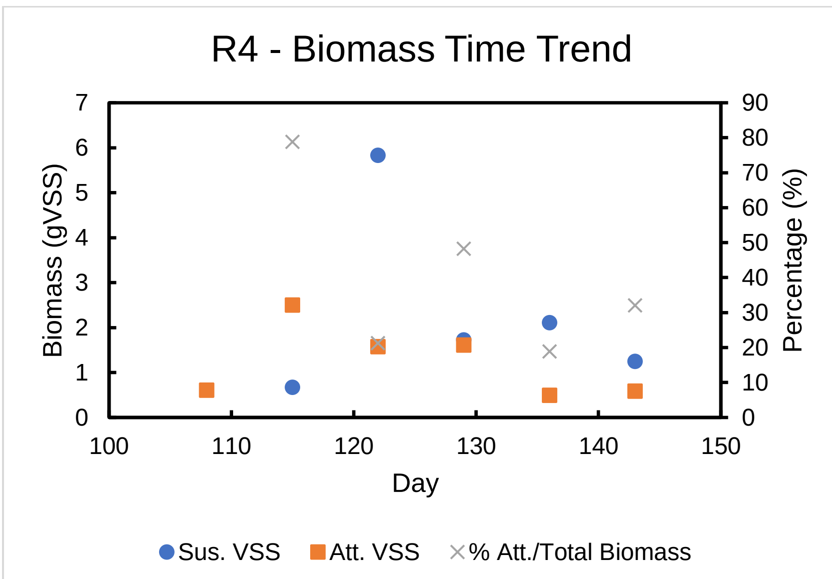

Figure 5-17 Reactor 4 biomass time trend

5.3.6 Biomass Yield

The biomass yield was calculated for R2 and R3 by graphing the cumulative COD removed and VSS produced. The resulting slope of the data is the observed biomass yield. For R2 and R3, the yields were found to be 0.18 and 0.21 gVSS/gCOD. While lower than the typical observed yield of most attached growth (~0.25 gVSS/gCOD) and suspended processes (>0.3 gVSS/gCOD)(Metcalf and Eddy, 2003), the expanded clay's observed yield is among the highest observed for fluidized bed bioreactors. Many earlier studies showed yields between 0.1-0.15 gVSS/gCOD, and some studies have shown yields below 0.1 gVSS/gCOD. R1 and R4 showed yields of 0.27 and 0.4 gVSS/gCOD, which are on par with suspended growth processes.. The cumulative COD removal and VSS production charts are shown in Appendix 1.

This high observed biomass yield coupled with somewhat high effluent VSS indicates that the systems were losing biomass at too high of a rate to manage higher loadings. This would limit the maximum COD and nitrogen loadings that the expanded clay could handle as an FBBR carrier material.

5.3.7 Nitrogen Mass Balances

Mass balances were calculated for the nitrogen concentrations for each column. Nitrogen assimilation into biomass was estimated using the assumption of 0.12 gN/gVSS(Metcalf and Eddy, 2003) and applying the biomass yields to calculate VSS production. Table 10 shows the nitrogen balance for Phase 2. As alkalinity could not be measured during phase 1, a nitrogen alkalinity balances could only be conducted for Phase 2. Alkalinity balances for all four columns closed within 40% error from the theoretical and actual net change in alkalinity.

Table 5-10 Nitrogen and Alkalinity Balance

		Phase 2			
		R1	R2	R3	R4
Influent N-NH_4	g/d	0.72	0.72	0.72	0.72
Effluent N-NH_4	g/d	0.22	0.01	0.02	0.28
Nitrogen Assimilated[1]	g/d	0.37	0.19	0.25	0.28
Ammonia Nitrified[2]	g/d	0.13	0.52	0.45	0.16
NOx Denitrified[3]	g/d	0.00	0.04	0.05	0.00
Influent Alkalinity	g/d	8.89	8.89	8.89	8.89
Effluent Alkalinity	g/d	8.26	6.75	6.69	8.18
Th Alkalinity Consumed[4]	g/d	0.89	3.67	3.21	1.15
Th Alkalinity Recovered[5]	g/d	0.00	0.14	0.20	0.01
Th Net Alkalinity	g/d	0.89	3.53	3.02	1.14
Actual Net Alkalinity	g/d	0.63	2.14	2.20	0.71
% Error - Alkalinity	%	29	39	27	38

[1] Nitrogen Assimilated = COD removed (g/d)*Biomass yield*(gVSS/gCOD)*0.12gN/gVSS

[2]Ammonia nitrified = Inf Amm. (g/d) – Eff Amm. (g/d) – N Assim. (g/d)

[3]NOx Dentrified = Amm. Nit (g/d) – Eff. NOx (g/d)

[4]Alkanility Consumed = Amm Nit (gN/d)*7.07 gAlk/gN

[5]Alkalinity Recovered = Amm Denit (gN/d)*3.53 gAlk/gN

5.3.8 Statistical Analysis of Nitrogen Removal Performance

Since all four columns operated at different DO concentrations and had different amounts/ratios of attached to suspended growth, a multi-parameter regression analysis was performed using Microsoft Excel. The values of **Table 5-11** represent the variables used for the regression. The ammonia removal as gN/d was the y-variable and the DO and Sus/Att growth ratio were the x-variables. The regression plotted in **Figures 5-18 and 5-19** yielded an R-Square value of 0.67 and the residuals for the predicted ammonia removal rates, shown in **Table 5-11**, were within ± 0.1 or ± 20%, implying that there is a causality between the ammonia removal rates and the DO and biomass ratios. The respective coefficients for DO and the suspended to attached growth ratio were 0.060 and -0.0061, indicating that DO has far greater contribution (10x) than the biomass distribution to the nitrogen removal.

Table 5-11 Summary of regression inputs and outputs

Column	N-NH_4 rem	Sus/att	DO	Pred. N-NH_4 rem
	g/d	mg/mg	mg/L	g/d
R1	0.50	5.0	1.6	0.49
R2	0.71	2.2	5.6	0.74
R3	0.70	1.4	3.1	0.60
R4	0.45	1.6	2	0.53

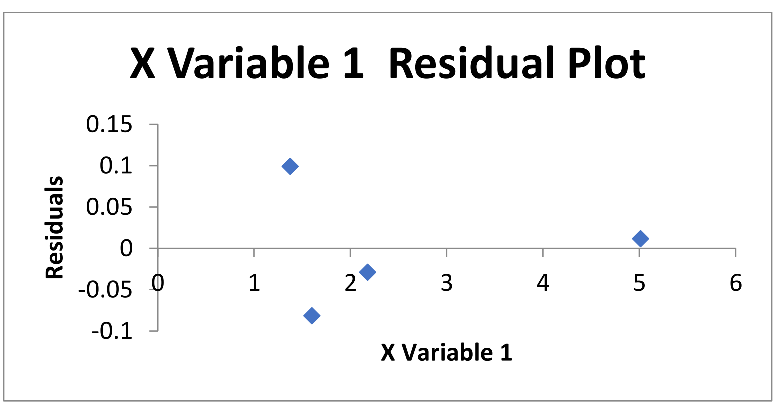

Figure 5-18 Suspended/attached VSS residual plot

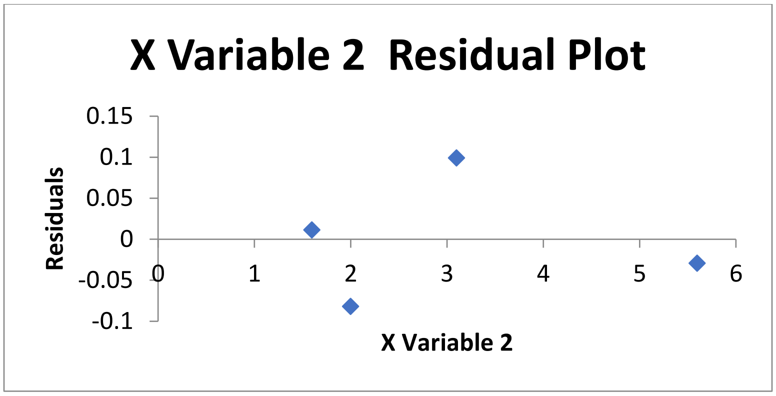

Figure 5-19 Dissolved oxygen residual plot

5.3.9 Comparison to Other Carriers

The results of studies on fluidized bed bioreactors were collected for comparison to the results of this study. For this purpose, aerobic single-column and two-column processes designed for nitrification or full nitrogen removal were used. Four single-column nitrifying FBBR studies and eight two-column CFBBR studies were reviewed. As the FBBR's of this study did not have dedicated anoxic columns for denitrification, only the nitrification/ammonia removal performance is evaluated between the studies. These included both upright and inverse fluidized bed bioreactor applications employing several different carrier materials.

From **Table 5-2**, it is observed that the nitrification dedicated studies were run at significantly higher NLR's compared the BOD and TN removal systems. Furthermore, their theoretical COD (ThCOD) were also significantly higher. It is also observed that the nitrification dedicated studies had considerably higher nitrogen removal by daily mass flow compared to the BOD and TN removal systems. It is observed that above 10 kgN/m^3/d the removal efficiency begins to decline dropping to 70% or lower, indicting a maximum limit of FBBR's for nitrogen removal(Tokutomi et al., 2010; H. Wang et al., 2019). These very high nitrogen loadings were treated without organics (COD) present or very low organic concentrations in the wastewater and so the nitrifiers would not face limitations due to competition with aerobic organic oxidizing bacteria(Metcalf and Eddy, 2003).

When comparing the COD and nitrification/nitrogen removal systems, we see that the expanded clay is more comparative to previous systems studied. The maximum loading is clearly higher for the upright CFBBR systems, which ranged from 2.0 to 8.5 kgThCOD/m^3/d(Andalib et al., 2010a; N. Chowdhury et al., 2008), while the IFBBR loadings were between 2.0 to 3.1 kgThCOD/m^3/d(H. Wang et al., 2020; L. Wang et al., 2021). As the upright applications used lava rock as carrier, its surface area is far higher than the plastic particles used in the inverse fluidized bed applications. However, it was also observed that the polypropylene-based system treating septic effluent showed much lower loadings of 1.3, 0.5 and 2.0 kg/m^3/d for COD, TN and ThCOD, respectively, compared to the other upright fluidized bed systems. In that study(Liu et al., 2019), the loadings had to be reduced due to very high ammonia concentrations, however, it still indicates that plastic carrier have lower performance quality compared to mineral-type carriers like lava rock.

However, it is observed that expanded clay reached similar loadings for COD, TN and ThCOD loadings compared to the other IFBBR studies, with expanded clay's highest ThCOD loading being 3.2 kgCOD/m-3-d and the highest of the IFBBR being 3.1 kgCOD/m^3/d. The removal rates for COD and N-NH_4 were also similar for the expanded clay and the other IFBBR studies, with both showing removals above 90% for both COD and N-NH_4.

Moving bed bioreactors were also compared as a process of similar design and operation. From Table 3, most of the MBBR application operated at fairly low OLR's (<1 kgCOD/m^3/d to 6 kgCOD/m^3/d) compared to the past CFBBR applications. The highest loadings shown for the MBBR were 6.25 and 4.7 kgCOD/m^3/d operating at HRT's of 8 and 24 hours, respectively. These two systems were treating high-strength and/or toxic wastewaters from the dairy and laundry industries(Bering et al., 2018; Santos et al., 2020). In terms of municipal wastewater treatment, the loadings shown for the MBBR did not go above 1.3 kgCOD/m^3/d, except for an application treating domestic greywater at a loading of 2.9 kgCOD/m^3/d(Chrispim & Nolasco, 2017). The organic loading alone for this system was higher than the expanded clay's maximum loading of 2.2 kgCOD/m^3/d, but this grey water had very little nitrogen.

5.4 Operational Challenges and Considerations

For these clays to be reliable for long-term operation, modifications may need to be considered. Wang et al. (L. Wang et al., 2021) and Wang et al. (H. Wang et al., 2020) both attempted surface modifications of plastic particles to improve attachment by coating the plastics in mineral-type materials; zeolite and activated carbon were shown to be the most effective. The expanded clay could potentially be used in a similar manner. Another possibility could be to use the clay in a composite material to maintain buoyancy or apply a coating to the clay to decrease its permeability to water and prevent absorption and subsequent settling.

The high suspended biomass concentrations could be exploited to achieve enhanced biological phosphorus removal. Both the CFBBR(N. Chowdhury et al., 2010) and MBBR [US Patent 10280099] have been demonstrated to facilitate EBPR by moving the media between the anoxic and aerobic zones/columns, but this requires significant solids handling capabilities to continuously move media throughout the system. With a hybrid system, the bulk liquid, which is high in biomass could be circulated, while keeping the media in their respective columns. A hybrid system would also be able to decouple the suspended and attached growth SRT's allowing for long SRT's for nitrifiers and shorter SRT's for the PAO's.

Finally, both inverse liquid-solid fluidization and gas-liquid-solid fluidization have been explored as means to reduce energy consumption for the CFBBR. As the expanded clay was fluidized by aeration only, it presents an opportunity for energy demand reduction for the CFBBR, as liquid circulation was not needed to maintain fluidization. Given the potential energy savings, the expanded clay could be a viable option to reduce the energy demand of the CFBBR. Wang et al. (2020) estimated the inverse fluidized bed bioreactor could have an energy demand 25% to 65% that of an upright fluidized bed bioreactor due to a combination of reduced aeration and reduced or eliminated liquid pumping requirements(H. Wang et al., 2020).Conclusion

5.5 Conclusions

Expanded clay was shown to be a viable option as a carrier for inverse aerobic fluidized bed bioreactors for wastewater treatment. The material demonstrated comparable COD removal and nitrification performance to previous IFBBR studies treating similar loadings and achieving comparable COD and ammonia removal of 93% and 98%, respectively. The expanded clay was also shown to be fluidized by aeration only, reducing the need for liquid pumping in the system to achieve fluidization.

The major caveat with the expanded clay was that it did not operate as a strictly attached growth process. During phase 2, high levels of suspended were present, indicating that the system was operating as a hybrid suspended and attached growth process. The high biomass production was also shown by the high observed yields for each column compared to past CFBBR studies. Overall, while the system achieved comparable COD and nitrogen removal to past fluidized bed studies, this material cannot be considered a strictly attached growth carrier.

Several areas still need further study to evaluate the expanded clay (and other expanded natural materials) as carrier media. This includes using expanded clay in a two-column CFBBR platform to evaluates it complete nitrogen removal performance, modifying the particles to overcome operational challenges such as water absorption, and larger-scale studies to better quantify its hydrodynamic properties.

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Chapter 6

Bed Voidage Predictions for Inverse Liquid-Solid Fluidized Beds

Abstract

Inverse liquid-solid fluidized beds are of particular interest for their applications to biochemical and wastewater treatment processes. Bed voidage is a key parameter for the design and scale-up of fluidized bed bioreactors, with the most commonly used method for predicting bed voidage being the Richard-Zaki equation. In this study, the bed voidage was measured for four sizes of Styrofoam particles with diameter and density ranging from 0.8 mm – 1.13mm and 28 kg/m^3 – 638 kg/m^3 in a conventional inverse liquid-solid fluidized bed. The experimental values of bed voidage, measured by the pressure drop method, were compared against bed voidage values predicted by the Richardson-Zaki equation using three different models for calculating the index value 'n'. The three models used several different correlations from literature to calculate the index value 'n' and the particle terminal velocity. Predicted bed voidage values from all three models were compared against experimental values from this study and previous studies with similar particles, with errors within ± 16%, demonstrating the validity of these Richardson-Zaki models for predicting bed voidage of small diameter and low-density particles.

6.1 Introduction

Inverse liquid-solid fluidization refers to a two-phase system where dispersed light solid particles, whose density is less than the liquid density, are suspended by a downward fluid flow. Liquid-solid fluidization systems have been applied successfully in several applications such as biological processes, wastewater treatment, biochemical and petrochemical technology, and food processing(Zhu et al., 2000). When upward liquid-solid fluidized beds are compared to inverse liquid-solid fluidization, the inverse liquid-solid fluidization systems present some advantages in practical applications. For example, light solid particles are characterized by intensive random movement, improving the liquid-solid contact efficiency, and enhancing the system's mass and heat transfer. Furthermore, biofilm thicknesses in wastewater treatment can be efficiently controlled (Renganathan & Krishnaiah, 2003). The increasing demand for liquid-solid reactors for many applications in fields of food production, biochemical engineering, and wastewater treatment (i.e., particle-supported biofilm), has created considerable interest in improved understanding of hydrodynamic characteristics for inverse liquid-solid fluidization configurations Click or tap here to enter text. (Metcalf and Eddy, 2003).

Bed voidage is a significant parameter for the design, operation and scale-up of fluidized bed bioreactors. Fan et al. (Fan et al., 1982) first studied bed voidage in a conventional inverse liquid-solid fluidized bed and suggested three models for predicting bed voidage. The first model was based on the Richardson-Zaki equation (Richardson & Zaki, 1954), which correlated the bed voidage to the ratio of the superficial liquid velocity to the terminal particle velocity. The second model was based on the Wen and Fan (Wen & Fan, 1974)correlation using a drag force function (f), a ratio of the liquid drag force in an inverse multi-particle fluidization system to that in a single-particle system. In the third model, height of bed expansion was directly correlated to the operating conditions and physical properties such as particle size, particle density, and liquid velocity. Fan et al. (Fan et al., 1982) concluded that the Richardson-Zaki equation, with a modified bed voidage index n, provided the best fit with their experimental data. Khan and Richardson (Khan & Richardson, 1989) then modified and generalized the bed voidage index in Richardson-Zaki equation to be a function of Archimedes number and the ratio of particle to column diameter for upward liquid fluidized beds, shown as Equation 6-1.

$$\frac{4.8-n}{n-2.4} = 0.043\, Ar^{0.57}[1 - 1.24\left(\frac{dp}{D}\right)^{0.27}] \qquad (6\text{-}1)$$

The flow behavior of free-rising light particles was initially thought to be similar to free-falling heavy particles since the exerted forces (i.e., drag and net buoyancy) on a single spherical particle has the same driving forces but in opposing directions. However, Karamanev and Nikolov (Karamanev & Nikolov, 1992b) experimentally demonstrated that light particles do not obey Newton's law for free settling particles. It was instead observed that a light particle whose density is less than 300 kg/m^3 or whose Reynolds number based on terminal particle velocity (Re_t) is larger than 130 accelerates and settles at a constant velocity in a spiral trajectory. As a result, the drag coefficient of the light particle is deviated from the standard drag curve to be 0.95, which is higher than the values of the standard drag curve based on Newton's Law (i.e., approximately 0.44). They found that index 'n' of the bed voidage calculated by the Fan et al. correlation (Fan et al., 1982) (1982) had deviation with the experimental data. They used the Richardson-Zaki equation (Richardson & Zaki, 1954) to describe their experimental data of bed voidage as the following Equations 6-2 and 6-3a-d.

$$\frac{U_l}{U_t} = \varepsilon^n \tag{6-2}$$

where n is the Richardson-Zaki index, U_t is the particle terminal velocity, and U_l is the superficial liquid velocity through the bed column. The value of the index, n, depends on the flow region of liquid-solid fluidization, where 'n' is constant at Stokes region and is a function of the Reynolds number (Re_t), and the ratio of the diameters of the particle and the column (d_p/D) in the transition region and is equal to 2.4 in the Newton region as showing in equations 6-2a-d:

$$n = 4.65 \qquad Re_t < 0.2 \tag{6-3a}$$

$$n = (4.4 + 18\frac{d_p}{D})Re_t^{-0.1} \qquad 1 < Re_t < 200 \tag{6-3b}$$

$$n = 4.4Re_t^{-0.1} \qquad 200 < Re_t < 500 \tag{6-3c}$$

$$n = 2.4 \qquad Re_t > 500\ . \tag{6-3d}$$

Karamanev (Karamanev, 1996) proposed a correlation in 1996 to estimate the drag coefficient for free-rising spherical particles, which is a function of the Archimedes number (Ar) with two conditions as shown in Equation 6-4a-b. This correlation is accurate for inverse conventional liquid-solid fluidization since the Ar number is defined based on the ratio of the difference between the gravitational and buoyancy forces and the viscous force, which are exerted on the suspended particle in a conventional fluidization regime.

$$C_D = 0.95 \qquad \text{when} \quad Ar > 1.8 \times 10^6 d_p^2 \tag{6-4a}$$

$$C_D = \left[\frac{432}{Ar}\left(1 + 0.047\ Ar^{\frac{2}{3}}\right) + \frac{0.517}{1+154\ Ar^{-\frac{1}{3}}}\right] \qquad \text{when} \quad Ar < 1.8 \times 10^6 d_p^2 \tag{6-4b}$$

Calderon et al. (Garcia-Calderon et al., 1998b) measured bed voidage of an inverse liquid-solid fluidized bed bioreactor using low density particles (213 kg/m^3) and a column with 0.08 m ID and 1 m in height. Their bed voidage data was compared to different models for predicting bed voidage for upward and inverse liquid-solid fluidized beds, and only agreed with Richardson-Zaki equation after substantiating 40% of the values of terminal particle velocity which found from standard drag curve. This could be because the drag coefficients of free-rising particles are typically larger than that of free-settling particles as proposed by Karamanev and Niklov (Karamanev & Nikolov, 1992a). Ulaganathan and Krishnaiah (Ulaganathan & Krishnaiah, 1996) proposed a different empirical correlation to predict bed voidage of inverse liquid-solid fluidized beds when using large particle diameters 20, 12.5, 12.5, and 20 mm and low densities 126, 216, 380, and 534 kg/m^3.

Lee (Lee, 2001) found good agreement between experimental bed voidage data of inverse liquid-solid fluidization and the Richardson-Zaki equation for solid particles with densities which are much closer to the liquid density (water). Yang and Renken (J. Yang & Renken, 2003) developed a generalized correlation based on equilibrium forces on a single suspended-particle for the upward liquid-solid fluidized bed. However, this correlation is more complicated because extra parameters have been added and the correlation is only governing the upward liquid-solid fluidized bed. Brown and Lawler (Brown & Lawler, 2003) analyzed most of experimental data of previous works for free-settling velocity of a spherical particle. They concluded that Turton and Clark's (Turton & Clark, 1987) correlation (Equation 6-5) provides high accuracy, within 2.5%, predicting free-settling velocities of a spherical particle compared to other knowing correlations.

$$Re_t = \frac{d_p U_t \rho_l}{\mu_l} = Ar^{1/3}[(\frac{18}{Ar^{\frac{2}{3}}})^{0.824} + (\frac{0.321}{Ar^{\frac{1}{3}}})^{0.412}]^{-1.214} \quad (6\text{-}5)$$

Renganathan and Krishnaiah (Renganathan & Krishnaiah, 2005) used the drift velocity model of Wallis (Wallis, 1969) to predict bed voidage of inverse liquid-solid fluidization, and it generally agrees with the Richardson-Zaki equation, which both depend on the relation between relative velocity of phases and bed voidage. Andalib et al. (Andalib, Zhu, et al., 2012) provided a new definition for bed voidage index in the Richardson Zaki equation to predict bed voidage of biofilm-coated particles in an anaerobic biological fluidized bed based only on the Ar number, instead of Re number. Das et al. (Das et al., 2015)determined an empirical correlation for bed expansion of inverse liquid fluidized beds for a variety of polymeric particles with varying density (900, 915, 919, and 944 kg/m3) and four different non-Newtonian fluids. This correlation is a function of static bed height, Re and particle and column diameter ratio, but it was developed only for non-Newtonian liquids (solutions of sodium salt of cellulose).

The Richardson-Zaki equation is still a popular correlation used to predict bed voidage for upward and inverse liquid-solid fluidized beds; however, in inverse liquid-solid fluidized beds, the literature review shows that some authors have proposed their empirical correlations or modifying the terminal particle velocities and/or the bed voidage index in the Richardson-Zaki equation to predict their experimental results of bed voidage. All the experimental measurements have been done in a large-scale system with a downer column bioreactor of 200 mm internal diameter and a height of 4.5 m to minimize wall effects. In order to govern the transition and Newton region in conventional inverse liquid-solid fluidized bed, some experimental data of previous studies Karamanev and Nikolov (Karamanev & Nikolov, 1992b) and Lee (Lee, 2001)(Lee, 2001)(Lee, 2001)(Lee, 2001)) are also used in this study. Experimental data from this study and previous studies were compared to the Richardson-Zaki equation with the terminal particle velocity calculated by the drag coefficient formula of Karamanev (Karamanev & Nikolov, 1992a), bed voidage index calculated by Khan and Richardson (Khan & Richardson, 1989) and calculating the terminal particle velocity directly from Brown and Lawler (Brown & Lawler, 2003) correlation. The solid particles properties in some previous works were summarized in **Table 6-1**.

Table 6-1 Solid particles used in published works in literature

Author	ρ_p $[\frac{kg}{m^3}]$	d_p $[mm]$	Re_t	$Ar \times 10^5$	System
(Fan et al., 1982)	930 882 887 822 388	6.35 9.53 19.1 9.53 10	- - - - -	2 13 100 20 76	Inverse liquid-solid fluidized bed, all solid particles located in Newton region where Ar $> 1.78125 \times 10^5$
(Garcia-Calderon et al., 1998b)	213	0.968	231	-	Inverse liquid-solid fluidized bed, but the particles properties are not irregular and not shape homogeneous
(Ulaganathan & Krishnaiah, 1996)	126 216 380 534	20 12.5 12.5 12.9	- - - -	- - - -	Inverse liquid-solid fluidized bed, For all particles located at Newton region, $10^6 < Ar < 7 \times 10^6$ And $512 < Re_t < 2040$
(J. Yang & Renken, 2003)	8800	0.135	4.34	0.0011	Upward liquid-solid fluidized bed
(Brown & Lawler, 2003)	-	-	-	-	This study for free-settling particle (previous data)
(Renganathan & Krishnaiah, 2005)	250 610 835 846 860	12.5 12.9 12.2 8 6.1	- - - - -	140 82.1 29.4 7.73 3.12	Inverse liquid-solid fluidized bed, All solid particles located at Newton region where Ar $>$ 1.78125×10^5

(Das et al., 2015)	915 919 944 900	5.64 (cylinder) (cylinder) (Disc)	- - - -	- - - -	Non-Newtonian inverse liquid-solid fluidized bed
(Karamanev & Nikolov, 1992b)	314 427 292 155 650 705 854 930 75 96 159 201	2.33 2.75 1.55 2.4 1.31 3.16 3.03 3.57 3.46 7.24 5.77 5.35	300 426 165 409 83 370 251 206 680 2350 1290 1400	1.08 1.17 0.26 1.15 0.08 0.91 0.4 0.32 3.76 33.9 15.8 12	Inverse liquid-solid fluidized bed
(Lee, 2001)	910 930 946	5.8 5.8 5.8	464 406 348	1.7 1.3 1.03	Upward and inverse liquid-solid fluidized bed

6.2 Materials and Methods

6.2.1 Experimental setup

A schematic of the experimental inverse liquid-solid fluidized bed system is shown in **Figure 6-1**. The Plexiglas downer column has an internal diameter of 0.2 m and a height of 4.5 m. The liquid distributor was installed at the top of the downer that has multiple pipes covered by a mesh to keep the particles in the column. The liquid (tap water) was pumped from the liquid reservoir through calibrated rotameters to the distributor at the top of the column and then returned at the bottom of the column to the reservoir. The liquid return pipe has the maximum height in the system before returning to the reservoir to ensure that the column downer does not have any air bubbles during experiments. Six pressure ports were installed at different heights of (84, 379, 734 1094, 1444, and 1794 mm) below the main liquid distributor, connecting to six manometers by tubes to measure the pressure profile across the fluidized bed. An optical-fiber probe was used to measure the voidage profiles in the downer column. The OFB measurements were calibrated using the experimental voidage measurements obtained from the pressure drop method. The open-end manometers were connected to a pressurized tank to control the water height by adjusting in tank pressure. Liquid flow rates were varied to cover a range of inverse liquid-solid fluidized bed operating conditions by rotameters. The rotameters were calibrated by measuring the volume of liquid flow leaving the column over a fixed period. All experiments were carried out at room temperature (25°) where the physical properties of water were taken. The physical properties of all types of spherical solid particles used in this study are listed in **Table 6-2**.

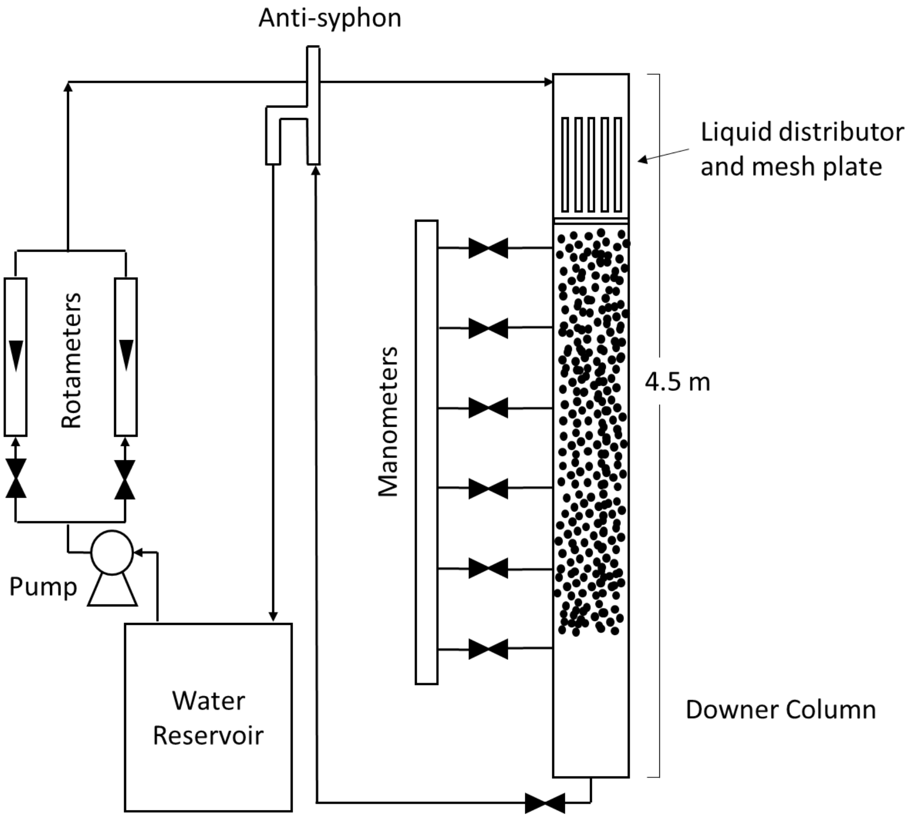

Figure 6-1 Schematic of inverse fluidized bed

6.2.2 Measurement techniques, calculations, models

Average solids holdup (solid phase) was determined by the pressure drop method, where the pressure drop along fluidized bed is measured by using manometers.

In the Richardson-Zaki equation, the parameter index "n" and the particle terminal velocity, U_t, can be experimentally determined by linearizing Equation 6-1 as follow Equation 6-6

$$Ln\ U_l = n\ ln(\varepsilon) + \ln U_t \tag{6-6}$$

When Equation 6-6 is plotted by using experimental data of U_l and ε for each type of solid particles, the index n is determined from the slope of Equation 6 and the terminal particle velocity of a single particle is defined at the bed voidage being equal to 1. In this study, the parameter "n" was obtained with the experimental data of bed voidage for solid particles with densities of 28, 122, 300 and

638 kg/m^3 which are denoted as ST028, ST122, ST300, and ST638, respectively. The experimental data from the studies of Karamanev and Lee were also used to compare with the results of this study.

Table 6-2 Solid particle properties for this study and previously studies

Particle material	Diameter [mm]	d_p/D	Density [kg/m^3]	Ar x 10^3	Authors
Styrofoam	0.8	0.004	28	6.1	This study
	1.13	0.006	122	15.6	
	1	0.005	300	8.6	
	1.1	0.005	638	5.9	
Styrofoam	2.33	0.029	314	108	(Karamanev & Nikolov, 1992b)
	2.75	0.034	426	117	
	1.55	0.019	165	26	
	2.4	0.030	409	115	
	1.31	0.016	83	8	
	3.16	0.040	370	91	
	3.03	0.038	251	40	
	3.57	0.045	206	32	
	3.46	0.043	75	376	
	7.24	0.091	96	3390	
	5.77	0.072	159	1580	
	5.35	0.067	201	1200	
Polyethylene	5.8	0.046	910	172	(Lee, 2001)
	5.8	0.046	930	134	
	5.8	0.046	946	103	

6.3 Results and discussion

6.3.1 Experimental Results

The bed voidage in a conventional inverse liquid-solid fluidized bed has been measured experimentally using the pressure drop method with varying superficial liquid velocity. The Reynolds number for each particle was determined by using the terminal particle velocity which was measured experimentally. It should be mentioned for this work and previous work that the operating ranges of studying bed voidage for conventional regimes did not include the Stokes region because it is difficult to find solid particles with properties that provide Reynolds number based on terminal particle velocity to be less than 0.2.

Figure 6-2 shows the experimental values of bed voidage as a function of superficial liquid velocity for ST028, ST122, ST300, and ST638. The observed bed voidage increases gradually with increasing liquid flowrate as greater drag force on the solid particles resulted in increase in bed voidage. However, it can be noted that ST028, ST122 and ST300 have similar trends and slopes compared to bed voidage curve of ST638. Generally, the particles with higher Archimedes numbers require higher superficial liquid velocity to be fluidized due to the greater net effective buoyancy force.

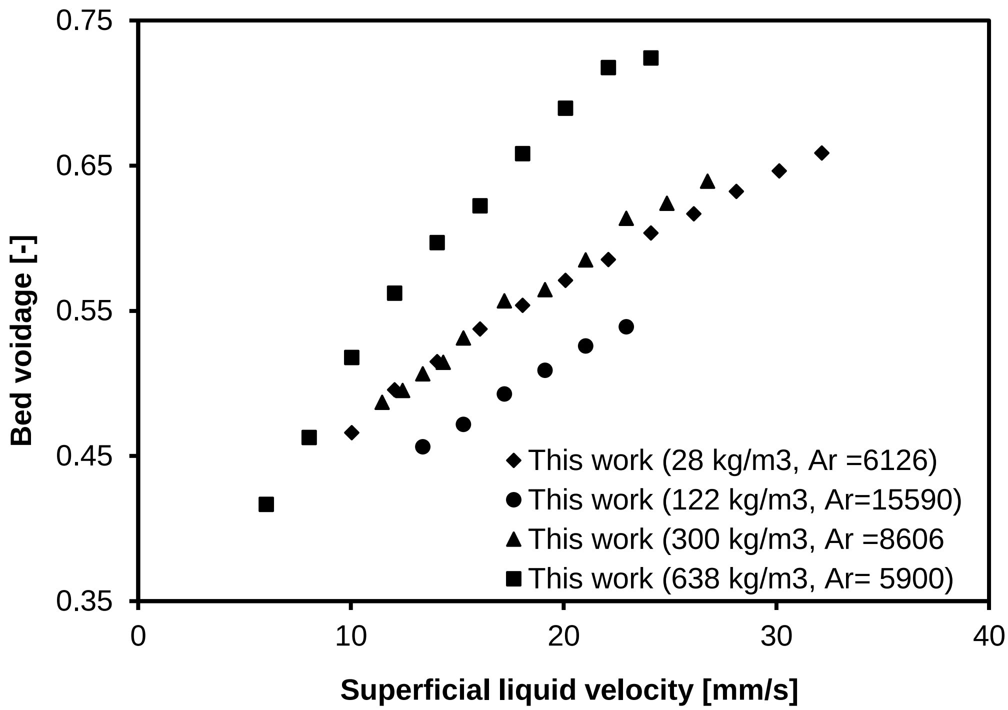

Figure 6-2 Variation of bed voidage as function of superficial liquid velocity

From **Figure 6-2**, as well as **Figure 6-3** below, it can be observed that the density, diameter, and the Archimedes number have the most significant effect on bed voidage. ST638, the particle with a density closest to that of water and the lowest Ar, showed the lowest solids holdup at lower liquid velocities. Whereas ST028, the lowest density particle, and ST122, the highest Ar, showed considerably higher solids holdup at the same liquid velocities. ST122 was shown to have higher solids holdups than ST028 at the same liquid velocities. This is due to ST122 higher particle diameter (1.1 mm) contributing to its considerably higher Ar number (15.6) than ST028 (6.1). As a result, ST122 requires the highest liquid velocity to achieve fluidization.

The linearized form of the Richardson-Zaki equation was used to determine the n-index and terminal velocity of the four particles. The slope of the linear trend produced gives the value for n and the intercept gives the terminal velocity. **Figure 6-3a-d** shows the linear trend of $\ln(U_l)$ and $\ln(\varepsilon)$ for ST028 (a), ST122 (b), ST300 (c), and ST638 (d). All four particles linear graphs showed high R^2 square indicating that trends are indeed linear. The values for ε and U_{lt} determined from the linearization are shown in **Table 6-3** along with experimental results from two other studies by Karamanev and Lee.

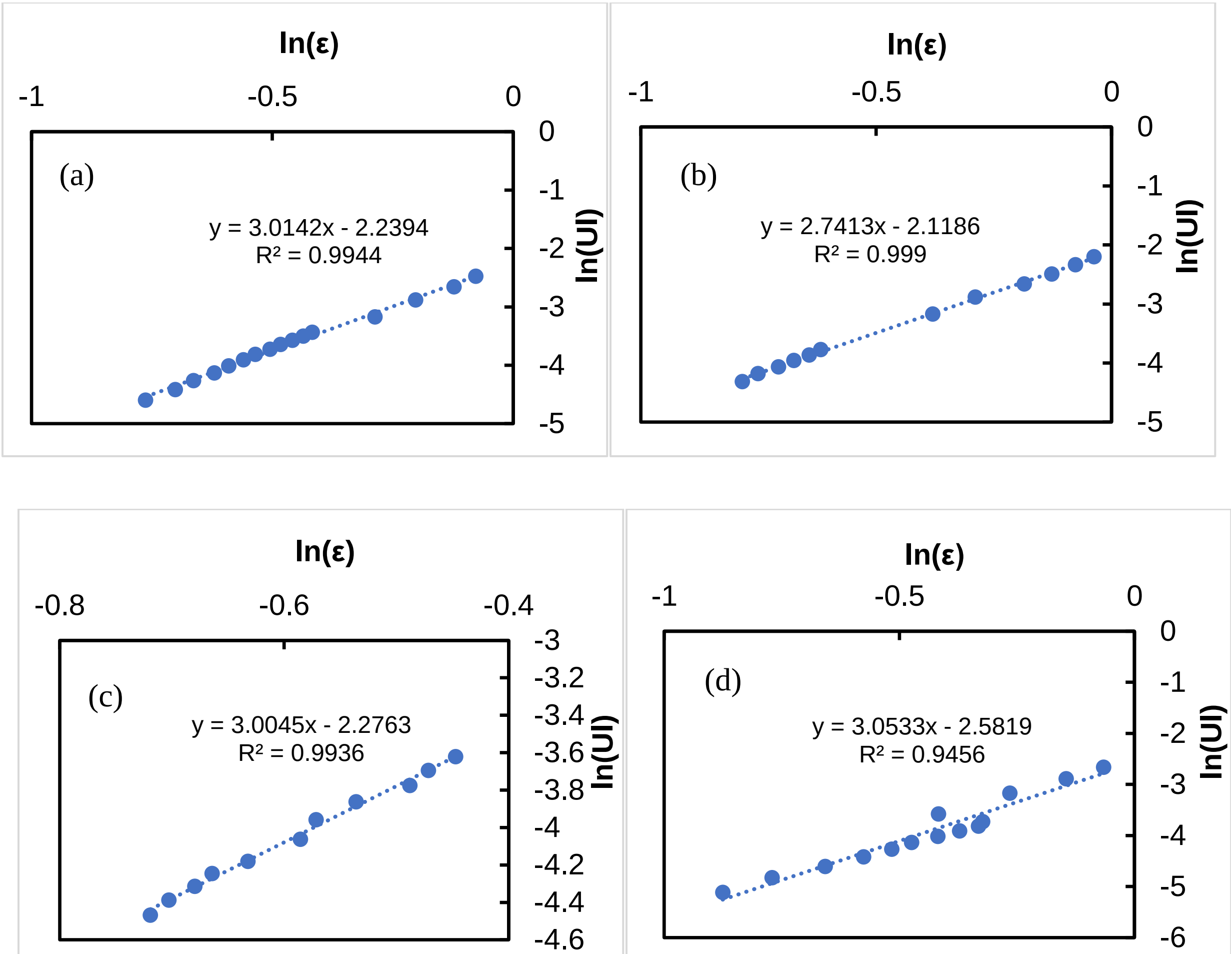

Figure 6-3a-d Linearization of liquid velocity and bed voidage for ST028 (a), ST122 (b), ST300 (c), and ST638 (d)

Table 6-3 Physical properties and experimental values of particles

	dp (mm)	Density (kg/m^3)	n (exp) (-)	Ult (exp) (m/s)	Ar (-)	Ret (-)
This Work	0.8	28	3.014	0.106	6126	95
	1.13	122	2.741	0.12	15590	151
	1	300	3.004	0.103	8606	115
	1.1	638	3.053	0.075	5900	92
Karamanev (1992)	2.33	314	2.54	0.12	108000	300
	2.75	427	2.5	0.143	117000	426
	1.55	292	2.66	0.102	26000	165
	2.4	155	2.59	0.159	115000	409
	1.31	650	3.25	0.061	8000	83
	3.16	705	2.57	0.107	91000	370
	3.03	854	2.71	0.076	40000	251
	3.57	930	2.77	0.052	32000	206
	5.77	159	2.51	0.026	1580000	1290

	3.46	75	2.34	0.021	376000	680
	5.35	201	2.44	0.225	1200000	1400
	7.24	96	2.41	0.264	3390000	2350
	5.8	910	2.41	0.08	172265	464
Lee (2001)	5.8	930	2.5	0.07	133984	406
	5.8	946	2.52	0.06	103360	348

6.3.2 Methods for Estimating Terminal Velocity and n-Index

With the data from this study, Karamanev and Lee several methods for estimating both bed voidage and terminal velocity were tested, comparing the theoretical predictions against the experimental results shown in Table 3. Newton's equations for the law of free-settling particles (Equation 6-7) using Karamanev's correlation for the drag coefficient (C_D) and the Turton and Clark correlation were tested for predicting terminal velocity. The Richard-Zaki and the Khan and Richardson correlation were used for predicting bed voidage. A simplified form of the Khan and Richardson correlation which neglects the wall effects ($d_p/D = 0$) was evaluated as well. The rationale behind this simplification is that the particles of very small diameter (~1mm) will have negligible wall effects present in the beds. The theoretical predictions are tabulated in **Table 6-4** with their corresponding particles and properties. Furthermore, the theoretical values are graphed and their corresponding experimental values in **Figures 6-4 and 6-5.**

$$U_t^2 = \frac{4g}{3}\frac{1}{C_D}\frac{d_{p(\rho_l-\rho_p)}}{\rho_l} \qquad (6\text{-}7)$$

Table 6-4 Predicted values of terminal velocity and n-index

	dp (mm)	Density (kg/m3)	Ut predictions		n predicitons		
			Newton LFS	Brown & Lawler	K&R	K&R-mod	R&Z
This Work	0.8	28	0.103	0.093	2.84	2.73	2.84
	1.13	122	0.117	0.119	2.69	2.61	2.72
	1	300	0.098	0.093	2.78	2.68	2.79
	1.1	638	0.074	0.066	2.86	2.74	2.86
Karamanev (1992)	2.33	314	0.148	0.179	2.54	2.47	2.49
	2.75	427	0.147	0.159	2.54	2.47	2.40
	1.55	292	0.123	0.118	2.67	2.56	2.86
	2.4	155	0.167	0.180	2.53	2.47	2.41
	1.31	650	0.079	0.067	2.85	2.69	3.01
	3.16	705	0.113	0.120	2.56	2.48	2.44
	3.03	854	0.077	0.078	2.64	2.53	2.53

	3.57	930	0.057	0.058	2.69	2.54	2.58
	5.77	159	0.258	0.304	2.44	2.42	2.40
	3.46	75	0.210	0.238	2.48	2.44	2.40
	5.35	201	0.243	0.284	2.45	2.42	2.40
	7.24	96	0.300	0.358	2.43	2.41	2.40
	5.8	910	0.083	0.093	2.52	2.46	2.38
Lee (2001)	5.8	930	0.073	0.081	2.54	2.46	2.41
	5.8	946	0.064	0.070	2.56	2.47	2.45

For the n-index predictions, all three methods fell within ±10% for most particles and values of n. It was observed that all three methods began to underpredict by a margin for experimental n values of 3 and above. The Khan and Richardson correlation had the lowest average absolute error at ±3.2%. The Richardson-Zaki method and modified Khan and Richardson correlation had average errors of ±4.3% and ±5.1%, respectively. However, it is observed in **Figure 6-4** that simplified Khan and Richardson correlation is more accurate for prediction of n values above 3. In the lower range of n values between 2.3 and 2.8, all three correlations showed relatively similar predictions, all within ±10%.

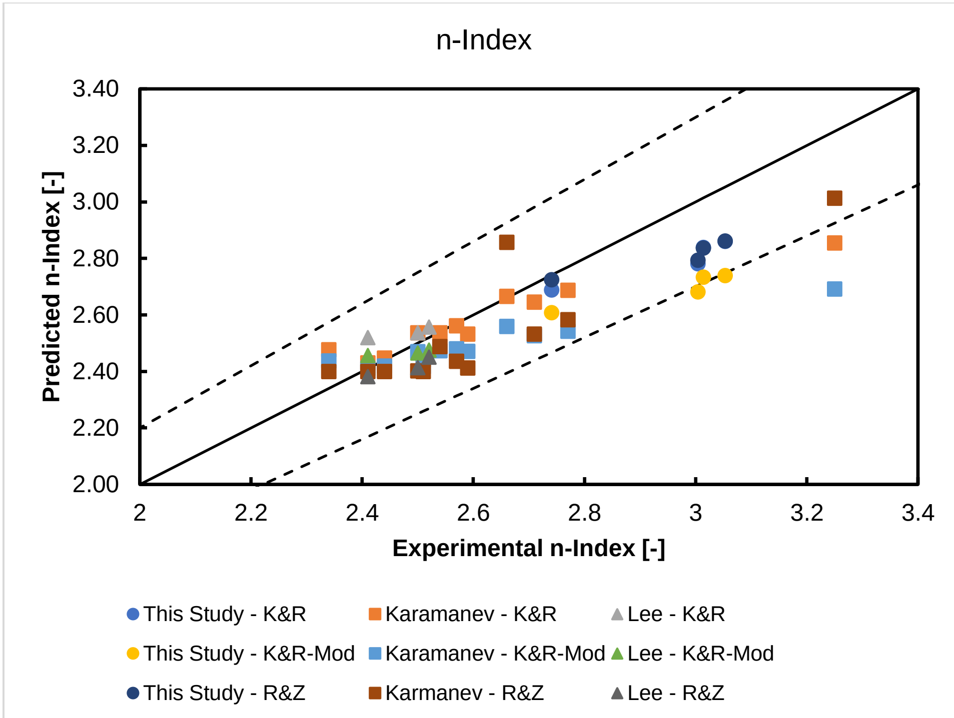

Figure 6-4 Comparison of experimental and predicted n-index values

Two methods for predicting particle terminal velocity were used; Newton's law of free settling particles with the drag coefficient calculated by Karamanev's correlation and Turton and Clark's correlation. Most predictions for both methods fell with ±20% of experimental values, however, the Newton-Karamanev method had an average error of 11%, while Turton and Clark showed 19% average error. Furthermore, it observed in **Figure 6-5** that error of the Newton-Karamanev method is significantly lower when for U_{lt} values above 0.16 m/s, with most Newton-Karamanev predicted values below 20% error and Turton and Clark above 20% error.

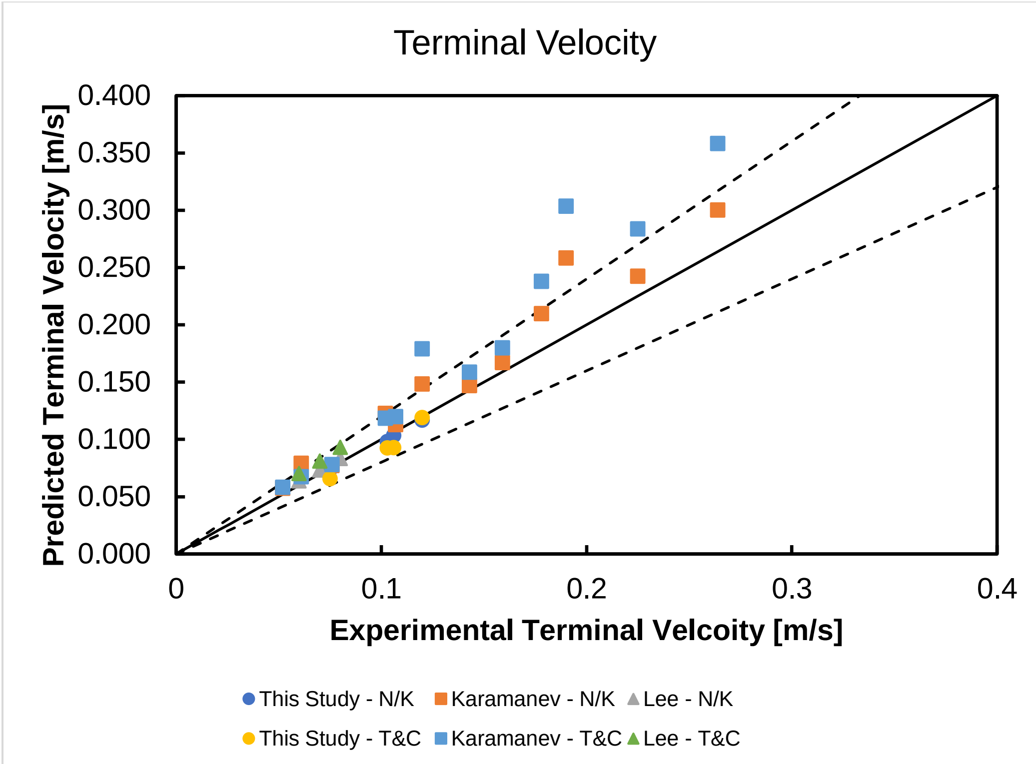

Figure 6-5 Comparison of experimental and predicted terminal velocity values

6.3.3 Suggested Approach

Terminal Velocity

From the results discussed above, it is clear that using the Newton equation with drag coefficients calculated by Karamanev's correlation is the superior method for estimating terminal velocity, with an average error of 10%. This was demonstrated for a wide range of particle properties; particle diameter from 0.8 mm to 7.2 mm and particle density of 28 kg/m^3 to 950 kg/m^3.

n-Index

For estimating the n-index, the methods explored yielded similar results with the linearized Richardson-Zaki equation and the modified and unmodified Khan and Richardson correlation giving predictions with average errors between 3% and 5%. However, it was observed that at higher values of the n-index (>3), the Richardson-Zaki equations were the most accurate with its predicted values.

6.4 Radial and Axial Solids Holdup Profiles

Figure 6-6a-d reports on the radial profiles of solids holdup for the four types of particles. For all types of particles, the radial distribution is fairly uniform but with a small increase towards the wall, due to the wall friction, which reduces the local liquid velocity at the wall boundary so more particles accumulate. This phenomenon is the same as in an upright liquid-solid fluidized bed (Ma et al., 2020; Zheng et al., 2002). Such nonuniformity is reduced by increasing the liquid velocity, which causes more turbulence and thus more horizontal movement of the particles. **Figure 6-6** also demonstrates that particles of low Ar require less liquid velocity for fluidization and achieving high voidage/low holdup. As seen in **Figures 6-6c and 6-6d** (ST300, Ar = 8600, ST638, Ar=5900), solids holdups well below 0.1 at 70 mm/s, whereas ST28 and ST122 had solids holdups between 0.1 and 0.15 at the same velocity.

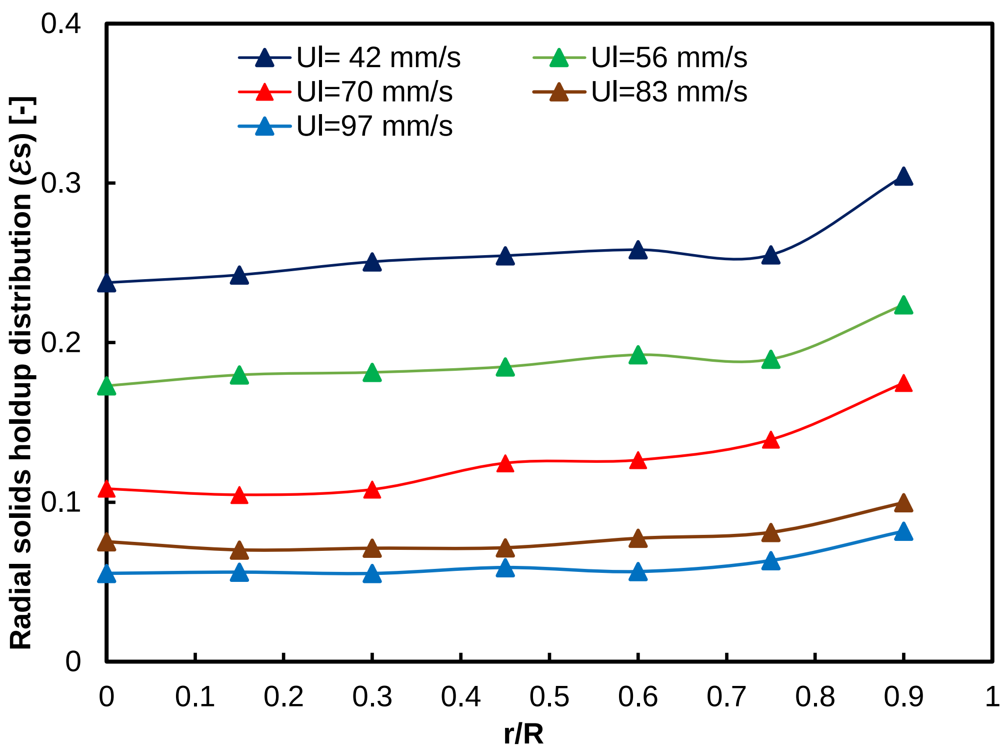

Figure 6-6a Radial solids holdup profiles ST028 at varying liquid velocities

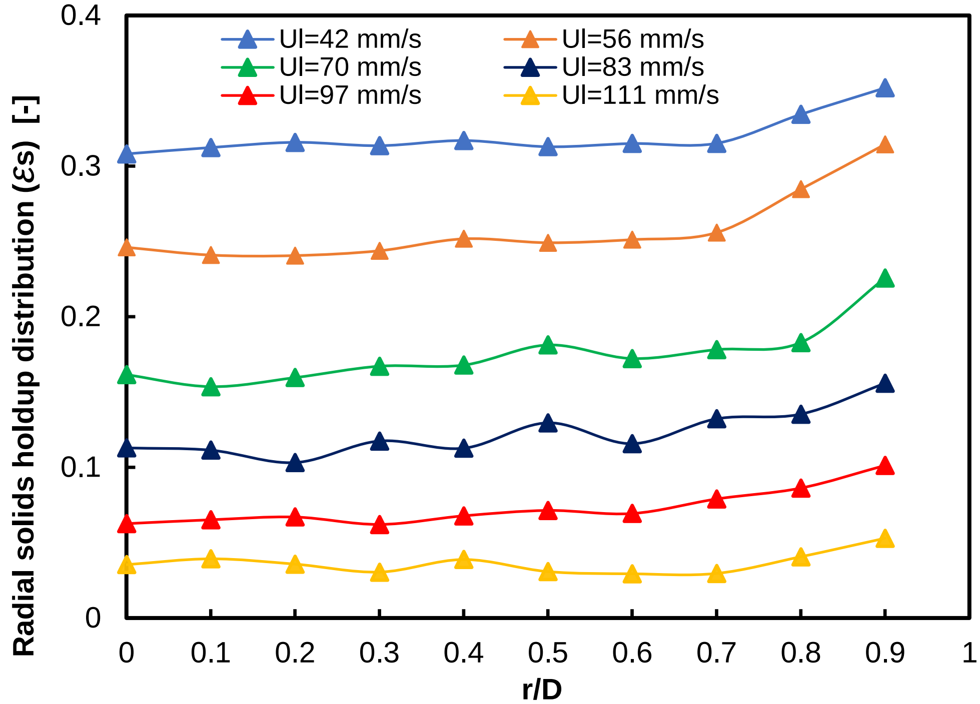

Figure 6-6b Radial solids holdup profiles ST122 at varying liquid velocities

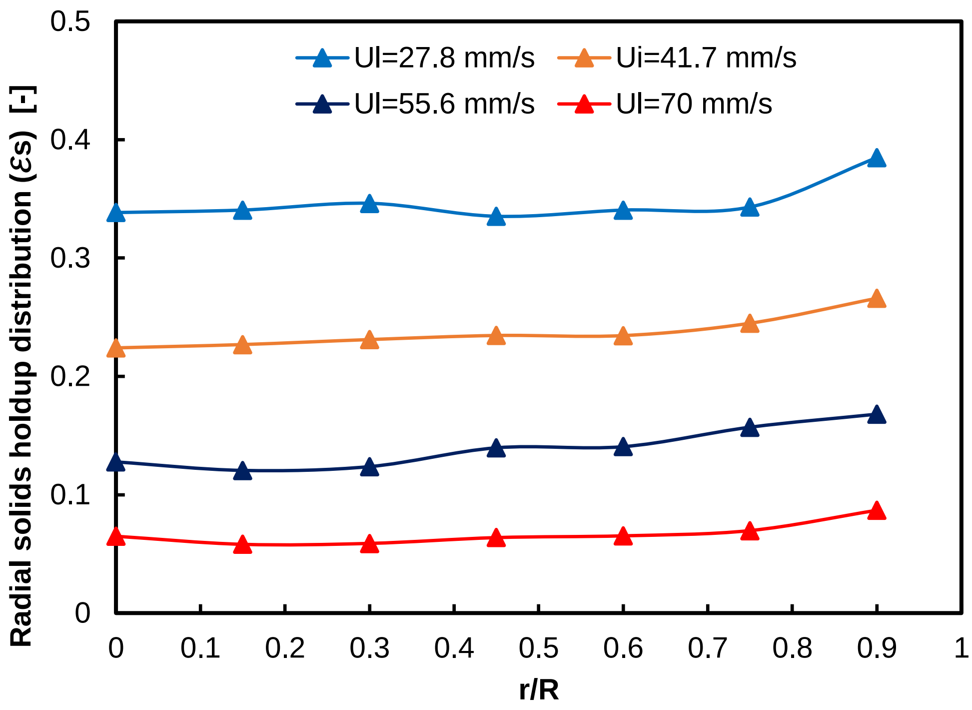

Figure 6-6c Radial solids holdup profiles ST638 at varying liquid velocities

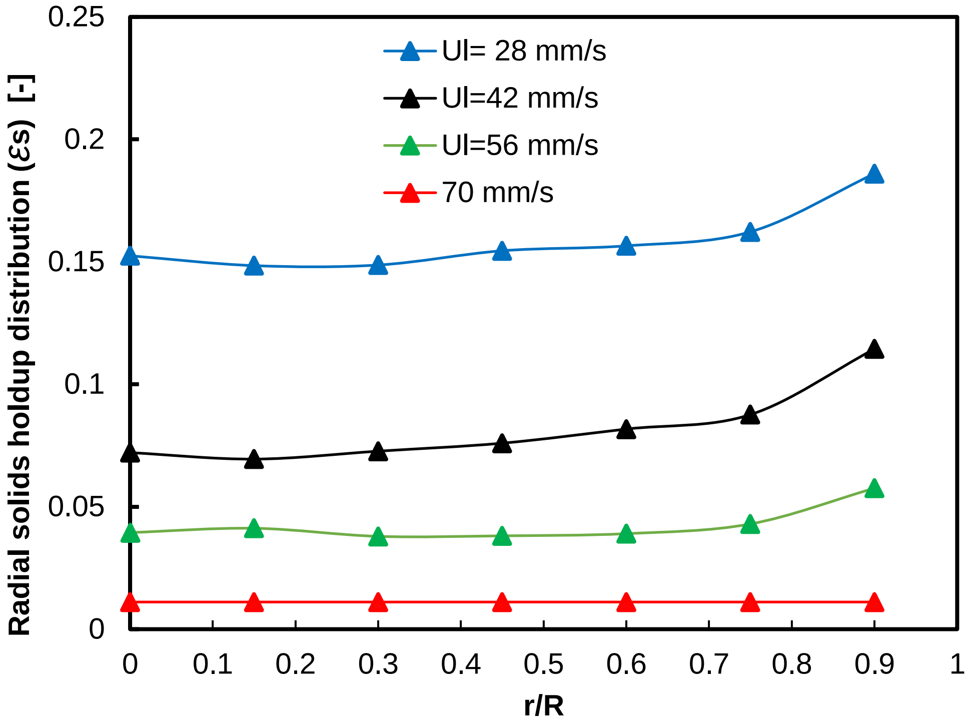

Figure 6-6d Radial solids holdup profiles ST300 at varying liquid velocities

The axial distributions of the solids holdup for ST028 particles are shown in **Figure 6-7**. At all velocities, the distribution is relatively uniform at the lower bed heights but drops off very quickly when reaching the top of the bed and exiting the dense phase. Like the wall effects discussed above, the near zero solids holdup observed above the dense phase is also observed in upright fluidization (Razzak et al., 2010; Song et al., 2019). It is also observed in **Figure 6-7** that the axial distribution becomes more uniform with increasing liquid velocity, due to the increase in turbulence and axial mixing in the column.

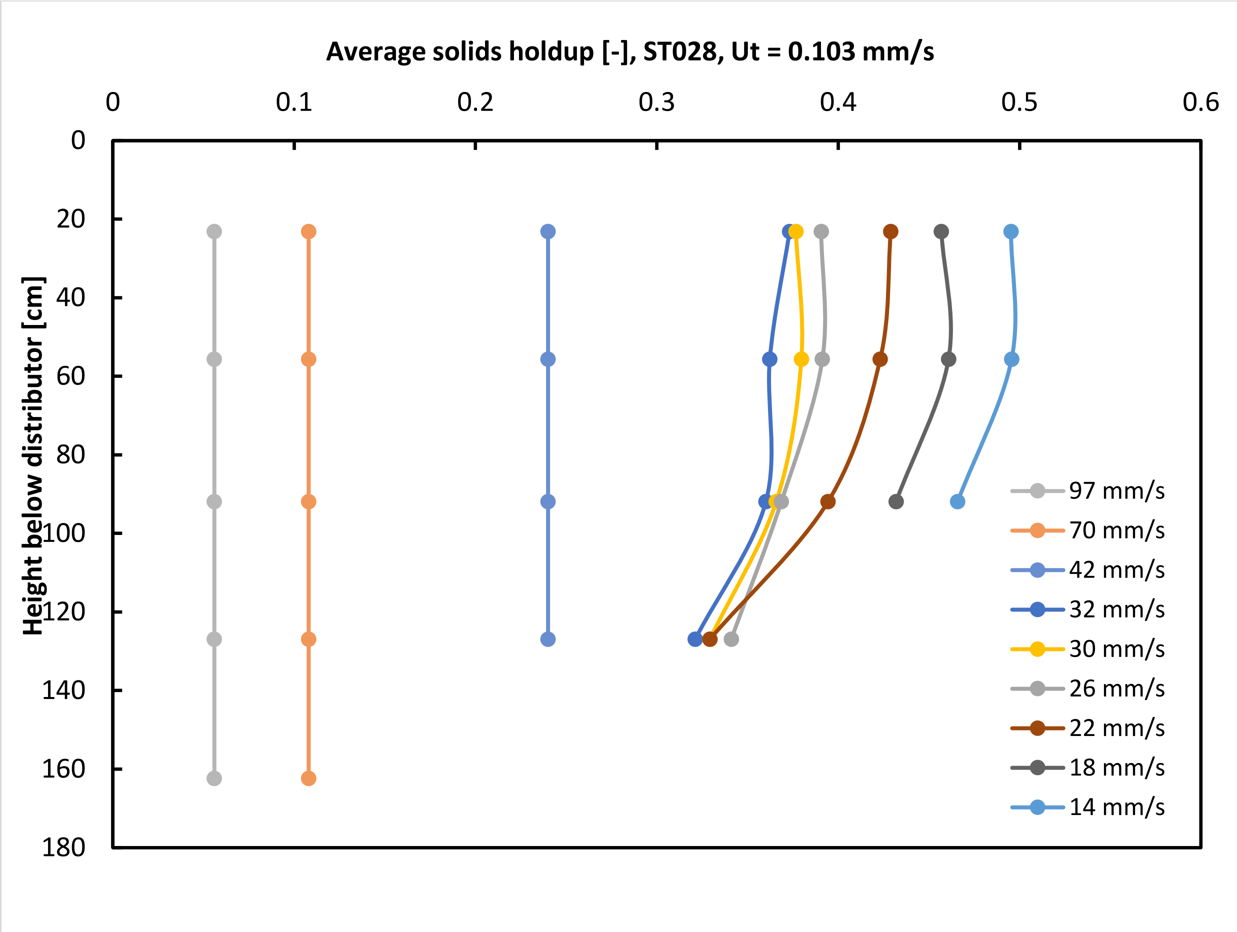

Figure 6-7 Axial solids holdup profiles ST028 at varying liquid velocities

The fluidized bed heights under different operating liquid velocities with the same amount of total particle inventory are shown in **Figure 6-8**. It is observed that increasing the fluidization (liquid) velocity would lead to a higher bed expansion and therefore higher bed voidage (lower solids holdup) and higher bed height (**Figure 6-8**). Furthermore, **Figure 6-8** reports the proportionality between the expanded bed height and voidage, as expected for fluidized beds, both inverse and upright.

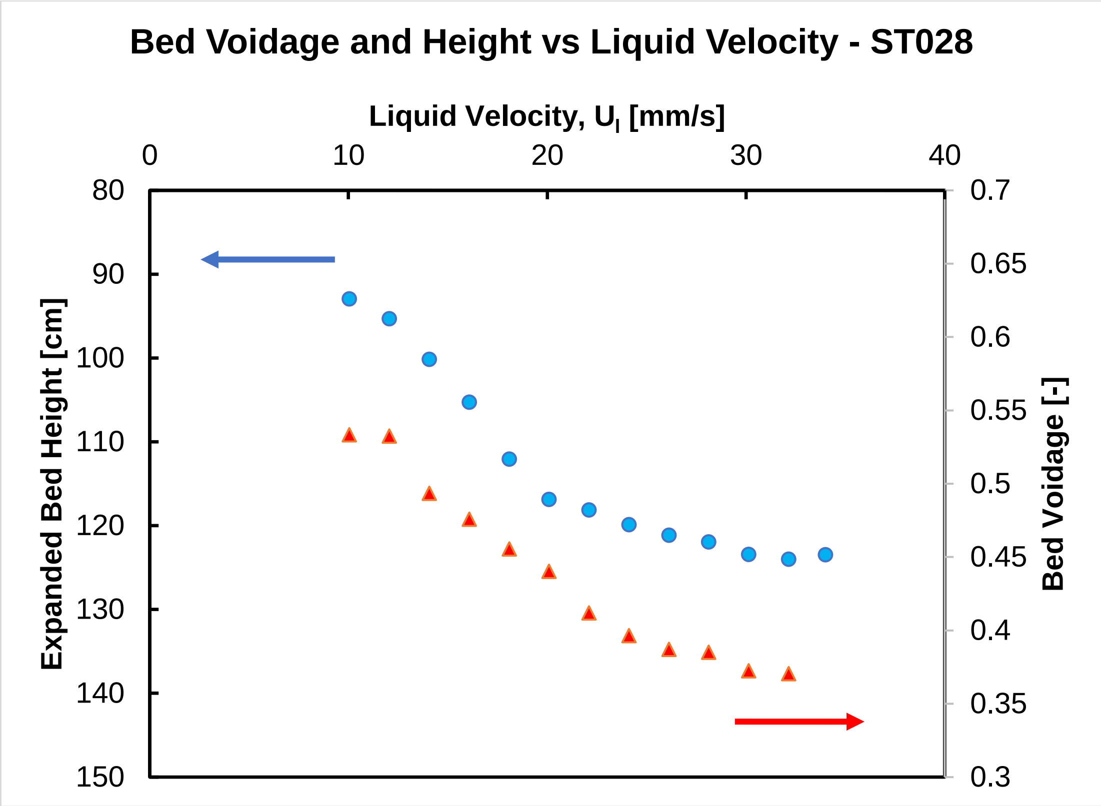

Figure 6-8 Expanded bed height and bed voidage of ST028 at varying liquid velocities

6.5 Conclusions

The average bed voidage was obtained by measuring the total pressure drop across the bed using a large column reactor as an inverse liquid-solid fluidized bed for four kinds of solid particles: ST028, ST122, ST300, and ST638. The linearized Richardson-Zaki was used to calculated the n-index and terminal velocity.

Several methods for estimating n-index and terminal velocity were evaluated using the particles from this study and two previous studies by Karamanev and Lee. The results of this analysis showed that Newton's equation for free settling particles using Karamanev's drag coefficient correlation is the most accurate for estimating terminal velocity with an average error of 10%. The three methods for estimating n-index were far closer in terms of average error. However, the Khan and Richardson correlation was the most accurate on average, but the Richardson-Zaki was the most accurate above n-index values of 3.

The radial solids holdup profiles for all four particles demonstrated that increasing Archimedes number and decreasing density generally contributes to higher liquid velocities required for inverse fluidization.

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Chapter 7

Summary and Conclusions

7.1 Summary

Chapter 3 evaluated the treatment of septic tank effluent in a pilot scale CFBBR. The pilot CFBBR used a polypropylene carrier for the treatment of the low COD:N ratio septic tanks effluent. The major findings were:

- 92%, 97%, and 82% COD, $N\text{-}NH_4$, and total nitrogen removal efficiencies were observed at OLR and NLR of 1.3 $kgCOD/m^3/d$ and 0.15 $kN/m^3/d$, respectively.
- The observed biomass yield was 0.12 gVSS/gCOD, consistent with previous CFBBR studies and significantly lower than conventional suspended and attached growth processes.
- The CFBBR achieved 50% reduction in supplemental carbon and 50% reduction in biosolids relative to a suspended growth process treating the same wastewater.

Chapter 4 used combined approach of using GPS-X and CapdetWorks to perform a cost-benefit analysis on the CFBBR against conventional treatment processes i.e. the MLE, the A2O and UCT processes, the four and five-stage BardenPho processes (4-BPD, 5-BPD), biological aerated filters (BAF), integrated fixed-film activated sludge (IFAS), and moving bed bioreactors (MBBR). The major findings were as follows:

- CFBBR had 10-20% lower Net Present Value based on CapEx and OpEx at low flows of 1 MGD to 5 MGD but similar costs (<10%) at flows of 10 MGD to 30 MGD.
- The major contributors to the CFBBRs NPV were the modularity of the CFBBR at higher flows, the liquid pumping energy for fluidization, and chemical usage for phosphorus removal.
- The CFBBR had 30% to 50% and 20% to 40% lower CapEx for the secondary treatment units and biosolids management units, respectively.

Chapter 5 tested multiple expanded mineral materials as biofilm carriers, focusing specifically on expanded clay after the start-up period as it did not present serious operational challenges. Four identical columns filled with clay at different fill fractions and aeration rates and configurations were operated as inverse fluidized bed bioreactors treating a synthetic wastewater with 350 mg/L COD and 35 mg/L $N\text{-}NH_4$.

- Three expanded mineral materials were tested as carriers for an inverse fluidized bed bioreactor; perlite, expanded glass bead, and expanded clay; only clay could operate effectively.
- At OLR and NLR of 2.2 and 0.2 kg/m^3/d, respectively, COD and $N\text{-}NH_4$ removal efficiencies of 93% and 98% were achieved.
- High levels of suspended biomass (>800 mg/L) and observed biomass yield above 0.19 gVSS/gCOD were measured, indicating that the clay could not sustain strictly attached growth operation.

Chapter 6 examines the Richardson-Zaki equation and newer empirical correlations for predicting bed expansion in inverse liquid-solid fluidized beds as a function of liquid velocity. Four sizes of Styrofoam particles tested in an inverse column, measuring solids holdup using the pressure drop method, as well as using an optical fibre-probe to measure the radial solids hold-up profiles. The results were used to calculate bed voidage and the linearized Richardson-Zaki equation was used to determine the experimental n-index and terminal velocity. Several correlations from literature were then compared in predicting the n-index and terminal velocity of the particles of this study and two previous studies of similar focus.

- For predicting the n-index, the Khan and Richardson correlation demonstrated ± 3% error, while the Richardson-Zaki correlations for ‘n’ had an average error of ± 5%, but were accurate above for n values 3.
- For predicting terminal velocity, Karamanev’s drag coefficient coupled with Newton’s terminal velocity equation for free settling particles showed error within ± 10%, superior to the Turton and Clark correlation with ± 20% error.

Table 7-1 Summary of Thesis Chapters, Major Results and Relevance

Chapter	Major Results	Relevance to thesis overall
Chapter 4 - The Circulating Fluidized Bed Bioreactor as a Biological Nutrient Removal Process for Municipal Wastewater Treatment: Process Modelling and Costing Analysis	-Identified small flow sizes (<5 MGD) as optimal for cost saving, according to NPV results -Major cost contributors are; fluidization energy, aeration energy, chemical phosphorus removal, and dewatering	-The CFBBR is only cost-competitive for a small range of rows currently, not suitable for large-scale applications due to system modularity and energy demand
Chapter 3 - Decentralized wastewater treatment in an urban setting: a pilot study of the circulating fluidized bed bioreactor treating septic tank effluent	-Communal wastewater treatment market in China is suitable for CFBBR platform, due to its low yield and subsequent low biosolids treatment costs -Though supplemental COD is required for TN removal in high-nitrogen wastewater, low biomass yield makes CFBBR superior to suspended growth processes	-CFBBR can meets the treatment needs of China's communal wastewater market and handle challenges related to high nitrogen concentrations while still remaining cost effective
Chapter 5 - Expanded mineral materials as carrier media in an inverse three-phase fluidized bed bioreactor for wastewater treatment	-Expanded proven capable of achieving nitrification at moderate nitrogen loading rates -Expanded clay did not achieve strictly attached growth treatment, as evidenced by high suspended biomass observed in bulk liquid	-Expanded clay has the potential to be used a carrier in hybrid suspended-attached growth process and be optimized for total nitrogen and phosphorus removal
Chapter 6 - Bed Voidage Predictions for Inverse Liquid-Solid Fluidized Beds	-Inverse fluidization has shown potential for reducing energy demand on the CFBBR -Several existing correlations are proven effective for estimating bed voidage characteristics of low density, small diameter particles	-With correlations that accurately estimate fluidization characteristics (bed voidage to corresponding liquid velocity), energy demand from fluidization can be estimated for system during the design stage and assess whether a particle's energy demand will contribute too much to the cost

7.2 Conclusions

The main goals of this work were to assess the capabilities of the CFBBR in terms of treatment performance for municipal wastewater treatment applications, evaluate its cost effectiveness on large-scale applications, and finally identify and explore aspects of the process requiring further development to optimize cost effectiveness. The major conclusions are as follows:

a. A pilot-scale CFBBR was tested for treating low COD:N nitrogen wastewater achieving biological nitrogen removal via nitrification and denitrification. The system was found to be COD limited for denitrification and required supplemental carbon to facilitate complete nitrogen removal. With the additional supplemental carbon, COD, N-NH_4, and TN removal rates of 92%, 97%, and 82%, respectively, were achieved. Further analysis of the observed biomass yield and biosolids generation showed the CFBBR, even with supplemental carbon, produces 60% less biosolids than conventional biological treatment processes (0.12 gVSS/gCOD vs 0.3 gVSS/gCOD), and so, would be an ideal process for decentralized and onsite residential wastewater treatment. Such a system would be compact, efficient, and require less chemicals usage and produce lower sludge quantities.

b. The cost analysis performed in Chapter 4 demonstrated the CFBBR being cost-effective for low flow centralized treatment plants (1.5 MGD – 5 MGD). At flows from 10 MGD – 30 MGD, the CFBBR was shown to be comparable in costs to conventional biological nutrient removal processes. Detailed costing estimates showed many advantages for the CFBBR possess, namely, small reactor sizes, elimination of primary clarifiers, reduced biosolids production, management and disposal, and reduced land footprint. The costing estimates also indicated that the major contributors to the cost of the CFBBR were its modularity (many small units), fluidization and aeration energy, and costs for chemical phosphorus removal (precipitation), and these were the factors hindering its cost effectiveness at flows above 10 MGD.

c. The exploration of expanded mineral materials as carriers did yield a suitable attached growth carrier, as most were faced by operational issues. Only the expanded clay achieved even moderate nutrient removal performance but could not sustain strictly attached growth operation, as high levels of biomass were observed suspended in the bulk liquid. Though not the desired outcome, the presence of attached growth and suspended offers opportunities to explore enhanced biological phosphorus removal, as the bulk liquid could be circulated between anaerobic and aerobic zones, while maintaining attached growth in each column for nitrification and denitrification. Furthermore, as the expanded clay could be fluidized by aeration alone, its use a carrier would save on energy cost by forgoing liquid pumping-induced fluidization. As shown in chapter 4, liquid pumping for fluidization was shown to account for 50% - 60% of the total energy demand of the CFBBR.

d. As inverse fluidization was identified in Chapter 4 as a potential means for saving cost and energy for the CFBBR, bed voidage characteristics of low-density Styrofoam particles were explored and evaluated against multiple correlations for predicting the n-index and terminal velocity. The Khan and Richardson correlation for predicting the n-index and Karamanev's drag coefficient correlation used with Newton's equation for free settling particles for predicting terminal velocity were found to the most accurate and reliable for low density and low Archimedes number particles.

7.3 Recommendations

From the findings of this research, particularly the cost analysis of Chapter 4, research in the future should focus on the following aspects:

a. The cost of chemicals for phosphorus removal was considerably higher for the CFBBR due to the high phosphorus loading to the system and lack of primary clarification. Enhanced biological phosphorus removal has been observed in the CFBBR but has yet to be optimized. Studies should be conducted to change the design of the anoxic and aerobic columns to achieve nitrification, denitrification and EBPR simultaneously. The results of Chapter 4 showed the EBPR capable process had drastically lower costs than chemical phosphorus removal, and so the same can be expected for an EBPR-capable CFBBR.
b. As the fluidization energy was identified as a major contributor to the OpEx of the CFBBR and inverse fluidization is a possible avenue to reduce this cost, a predictive model or correlation would provide the means to estimate the liquid and/or air flow requirements and calculate the energy demand for fluidization. Such studies should be conducted on both clean and biofilm-laden particles. The biofilm-laden particle studies should be conducted at different loadings to measure the fluidization properties of the same particles with different amounts of biofilm. As the biofilm size affects the total size and density of the particle, a model that accounts for this will be useful estimating energy demand changes brought on by long-term changes on the system loading.
c. Develop a scale-up model for the IFBBR to maximize module size of (I)-CFBBR as modularity of the CFBBR was shown to inflate the capital cost and hinder cost

competitiveness at high flows. To have a firm maximum size for the CFBBR will provide a hard limit for the flow size that the CFBBR can be practically used for.

d. A follow-up costing study in the same format as Chapter 4 should be conducted based on an I-CFBBR, designed and optimized for enhanced biological phosphorus removal and full biological nutrient removal. As these were identified as two of the major hinderances to the CFBBR's cost competitiveness, a costing study would quantify the costing changes these improvements would provide. Furthermore, the study can be expanded to evaluate other applications of the fluidized bed bioreactor, such biosolids treatment via anaerobic digestion. Anaerobic digestion of biosolids has been demonstrated in an anaerobic fluidized bed platform and showed superior treatment to suspended growth/CSTR digesters. To use fluidized bed digesters for biosolids treatment would further decrease the already lower capital costs for anaerobic digestion seen in Chapter 4 for the CFBBR-based treatment plants.

Appendices

Appendix 1 (Chapter 4)

GPS-X Modelling Results

Table A1.1 Influent and Effluent of GPS-X Model for CFBBR

		Phase 2				Phase 3			
		Influent		Effluent		Influent		Effluent	
Parameter	Units	Experimental	Model	Experimental	Model	Experimental	Model	Actual	Model
Flow	m^3/d	1.025	1.025	1.025	1.025	1.27	1.27	1.27	1.27
TSS	mg/L	43 + 24	49	17 + 10	18	30 + 6	30	32 + 17	29.4
VSS	mg/L	39 + 23	45	14 + 9	14	27 + 5	26	26 + 14	22.9
$cBOD_5$	mg/L	410 + 230	381	7 + 3	6	658 + 77	572	17 + 11	3.19
$scBOD_5$	mg/L	403 + 229	352	2 + 1	2	647 + 80	564	3 + 2	1
COD	mg/L	585 + 290	584	45 + 9	52	857 + 130	858	70 + 20	66.9
SCOD	mg/L	503 + 267	519	30 + 11	30	797 + 129	818	31 + 9	32.4
$N\text{-}NH_4$	mgN/L	72 + 36	72	5 + 7	0.2	82 + 17	82	2 + 3	0.06
$N\text{-}NO_3$	mgN/L	0.2 + 0.1	0.2	13 + 8	12.3	0.6 + 0.2	0.6	13 + 5	21
TKN	mgN/L	91.9	91.9	5.6	4.1	99	99	5	4.3
TN	mgN/L	92 + 46	92.1	21 + 16	16.6	100 + 19	100	18 + 10	25.5
Alkalinity	$mgCaCO_3/L$	358 + 135	358	62 + 28	35	486 + 84	257	109 + 45	45.3

COD and Solids Balance Tables

Table A1.2 1.5 MGD COD and Solids Balance

Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Influent									
COD (kg/d)	2,840	2,840	2,840	2,840	2,840	2,840	2,840	2,840	2,840
TSS (kg/d)	1,250	1,250	1,250	1,250	1,250	1,250	1,250	1,250	1,250
Influent + Side Stream Recycle									
COD (kg/d)	3,035	3,005	2,980	2,980	3,020	3,060	2,960	2,915	2,915
TSS (kg/d)	1,602	1,500	1,390	1,390	1,445	2,245	1,875	1,475	1,550
Primary Sludge									
COD (kg/d)	1,225	1,225	1,210	1,210	1,225	1,255	1,205	1,180	-
TSS (kg/d)	934	880	810	810	840	1,320	1,100	860	-
Secondary Influent									
COD in (kg/d)	1,825	1,790	1,780	1,775	1,790	1,815	1,760	1,725	2,916
Y (gVSS/gCOD)	0.33	0.35	0.28	0.28	0.28	0.20	0.25	0.23	0.18
WAS									
TSS (kg/d)	1,260	1,295	700	700	720	525	895	1,115	1,515
VSS (kg/d)	595	620	500	500	510	370	435	405	530
Thickened Biosolids									
TSS (kg/d)	1,080	1,110	600	600	615	450	770	960	1,300
VSS (kg/d)	510	530	430	430	435	315	370	345	455
Digested Biosolids									
TSS (kg/d)	1,375	1,355	860	860	885	1,160	1,315	1,295	1,050
VSS (kg/d)	520	515	465	465	475	495	450	425	205
Dewatered Biosolids									
TSS (kg/d)	1,310	1,285	820	820	800	1,105	1,250	1,235	1,045
VSS (kg/d)	495	490	445	445	425	470	430	402	200
Hauling and Landfill									
TSS (kg/d)	1,310	1,285	820	820	800	1,105	1,250	1,235	1,045
VSS (kg/d)	495	490	445	445	425	470	430	402	200

Table A1.3 4.6 MGD COD and Solids Balance

Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Influent									
COD (kg/d)	8,700	8,700	8,700	8,700	8,700	8,700	8,700	8,700	8,700
TSS (kg/d)	3,830	3,830	3,830	3,830	3,830	3,830	3,830	3,830	3,830
Influent + Side Stream Recycle									
COD (kg/d)	9,655	9,230	9,575	9,575	9,715	9,495	9,105	8,930	8,750
TSS (kg/d)	5,785	4,607	5,435	5,435	5,685	6,645	5,775	4,520	4,650
Primary Sludge									
COD (kg/d)	3,931	3,750	3,900	3,900	3,970	3,865	3,695	3,620	-
TSS (kg/d)	3,378	2,690	3,185	3,185	3,325	3,890	3,380	2,634	-
Secondary Influent									
COD in (kg/d)	5,710	5,485	5,680	5,200	5,770	5,615	5,385	5,315	8,748
Y (gVSS/gCOD)	0.34	0.35	0.30	0.33	0.30	0.33	0.25	0.23	0.19
WAS									
TSS (kg/d)	4,125	3,970	2,450	2,450	2,535	2,640	2,750	3,425	4,645
VSS (kg/d)	1,925	1,905	1,700	1,700	1,735	1,850	1,325	1,245	1,620
Thickened Biosolids									
TSS (kg/d)	3,540	3,405	2,100	2,100	2,180	2,270	2,345	2,915	3,995
VSS (kg/d)	1,650	1,635	1,455	1,455	1,490	1,590	1,135	1,060	1,395
Digested Biosolids									
TSS (kg/d)	4,770	4,165	3,245	3,240	3,385	3,940	4,030	3,975	3,220
VSS (kg/d)	1,765	1,580	1,700	1,710	1,750	1,815	1,385	1,295	625
Dewatered Biosolids									
TSS (kg/d)	4,525	3,950	3,080	3,080	3,050	3,740	3,840	3,780	3,185
VSS (kg/d)	1,675	1,500	1,625	1,625	1,575	1,725	1,315	1,230	620
Hauling and Landfill									
TSS (kg/d)	4,525	3,950	3,080	3,080	3,050	3,740	3,840	3,780	3,185
VSS (kg/d)	1,675	1,500	1,625	1,625	1,575	1,725	1,315	1,230	620

Table A1.4 10 MGD COD and Solids Balance

Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Influent									
COD (kg/d)	19,000	19,000	19,000	19,000	19,000	19,000	19,000	19,000	19,000
TSS (kg/d)	8,333	8,333	8,333	8,333	8,333	8,333	8,333	8,333	8,333
Influent + Side Stream Recycle									
COD (kg/d)	20,231	20,031	19,865	19,865	20,131	20,398	19,731	19,431	20,000
TSS (kg/d)	10,679	9,999	9,266	9,266	9,632	14,965	12,499	9,832	10,800
Primary Sludge									
COD (kg/d)	8,166	8,166	8,066	8,066	8,166	8,366	8,033	7,866	-
TSS (kg/d)	6,220	5,861	5,395	5,395	5,594	8,791	7,326	5,728	-
Secondary Influent									
COD in (kg/d)	12,165	11,932	11,865	11,832	11,932	12,099	11,732	11,499	20,412
Y (gVSS/gCOD)	0.33	0.35	0.28	0.28	0.28	0.20	0.25	0.23	0.18
WAS									
TSS (kg/d)	8,399	8,632	4,666	4,666	4,800	3,500	5,966	7,433	10,785
VSS (kg/d)	3,966	4,133	3,333	3,333	3,400	2,466	2,900	2,700	3,755
Thickened Biosolids									
TSS (kg/d)	7,199	7,399	4,000	4,000	4,100	3,000	5,133	6,399	9,255
VSS (kg/d)	3,400	3,533	2,866	2,866	2,900	2,100	2,466	2,300	3,220
Digested Biosolids									
TSS (kg/d)	9,166	9,032	5,733	5,733	5,899	7,733	8,766	8,632	7,495
VSS (kg/d)	3,466	3,433	3,100	3,100	3,166	3,300	3,000	2,833	1,445
Dewatered Biosolids									
TSS (kg/d)	8,732	8,566	5,466	5,466	5,333	7,366	8,333	8,233	7,410
VSS (kg/d)	3,300	3,266	2,966	2,966	2,833	3,133	2,866	2,680	1,430
Hauling and Landfill									
TSS (kg/d)	8,732	8,566	5,466	5,466	5,333	7,366	8,333	8,233	7,410
VSS (kg/d)	3,300	3,266	2,966	2,966	2,833	3,133	2,866	2,680	1,430

Table A1.5 20 MGD COD and Solids Balance

Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Influent									
COD (kg/d)	37,500	37,500	37,500	37,500	37,500	37,500	37,500	37,500	37,500
TSS (kg/d)	16,500	16,500	16,500	16,500	16,500	16,500	16,500	16,500	16,500
Influent + Side Stream Recycle									
COD (kg/d)	40,500	40,100	39,700	39,700	40,300	40,800	39,500	38,900	38,500
TSS (kg/d)	21,400	20,000	18,500	18,500	19,300	29,900	25,000	19,700	20,200
Primary Sludge									
COD (kg/d)	16,300	16,300	16,100	16,100	16,300	16,700	16,100	15,700	-
TSS (kg/d)	12,453	11,733	10,800	10,800	11,200	17,600	14,666	11,466	-
Secondary Influent									
COD in (kg/d)	24,300	23,866	23,733	23,666	23,866	24,199	23,466	23,000	38,000
Y (gVSS/gCOD)	0.33	0.35	0.28	0.28	0.28	0.20	0.25	0.23	0.18
WAS									
TSS (kg/d)	16,800	17,266	9,333	9,333	9,600	7,000	11,933	14,866	19,995
VSS (kg/d)	7,933	8,266	6,667	6,667	6,800	4,933	5,800	5,400	6,980
Thickened Biosolids									
TSS (kg/d)	14,400	14,800	8,000	8,000	8,200	6,000	10,266	12,800	17,165
VSS (kg/d)	6,800	7,066	5,733	5,733	5,800	4,200	4,933	4,600	5,590
Digested Biosolids									
TSS (kg/d)	18,333	18,066	11,466	11,466	11,800	15,466	17,533	17,266	13,870
VSS (kg/d)	6,933	6,866	6,200	6,200	6,333	6,600	6,000	5,667	2,690
Dewatered Biosolids									
TSS (kg/d)	17,466	17,133	10,933	10,933	10,666	14,733	16,666	16,466	13,700
VSS (kg/d)	6,600	6,533	5,933	5,933	5,667	6,267	5,733	5,360	2,660
Hauling and Landfill									
TSS (kg/d)	17,466	17,133	10,933	10,933	10,666	14,733	16,666	16,466	13,700
VSS (kg/d)	6,600	6,533	5,933	5,933	5,667	6,267	5,733	5,360	2,660

Table A1.6 30 MGD COD and Solids Balance

Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Influent									
COD (kg/d)	56,800	56,800	56,800	56,800	56,800	56,800	56,800	56,800	56,800
TSS (kg/d)	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000	25,000
Influent + Side Stream Recycle									
COD (kg/d)	60,700	60,100	59,600	59,600	60,400	61,200	59,200	58,300	58,300
TSS (kg/d)	32,040	30,000	27,800	27,800	28,900	44,900	37,500	29,500	31,000
Primary Sludge									
COD (kg/d)	24,500	24,500	24,200	24,200	24,500	25,100	24,100	23,600	-
TSS (kg/d)	18,680	17,600	16,200	16,200	16,800	26,400	22,000	17,200	-
Secondary Influent									
COD in (kg/d)	36,500	35,800	35,600	35,500	35,800	36,300	35,200	34,500	58,300
Y (gVSS/gCOD)	0.33	0.35	0.28	0.28	0.28	0.20	0.25	0.23	0.18
WAS									
TSS (kg/d)	25,200	25,900	14,000	14,000	14,400	10,500	17,900	22,300	30,700
VSS (kg/d)	11,900	12,400	10,000	10,000	10,200	7,400	8,700	8,100	10,715
Thickened Biosolids									
TSS (kg/d)	21,600	22,200	12,000	12,000	12,300	9,000	15,400	19,200	26,305
VSS (kg/d)	10,200	10,600	8,600	8,600	8,700	6,300	7,400	6,900	9,180
Digested Biosolids									
TSS (kg/d)	27,500	27,100	17,200	17,200	17,700	23,200	26,300	25,900	21,385
VSS (kg/d)	10,400	10,300	9,300	9,300	9,500	9,900	9,000	8,500	4,150
Dewatered Biosolids									
TSS (kg/d)	26,200	25,700	16,400	16,400	16,000	22,100	25,000	24,700	21,120
VSS (kg/d)	9,900	9,800	8,900	8,900	8,500	9,400	8,600	8,040	4,100
Hauling and Landfill									
TSS (kg/d)	26,200	25,700	16,400	16,400	16,000	22,100	25,000	24,700	21,120
VSS (kg/d)	9,900	9,800	8,900	8,900	8,500	9,400	8,600	8,040	4,100

Energy Demand, Chemical Dosage, and OpEx Benchmarks

Table A1.7 Energy Demand, Chemical Dosage, and OpEx Benchmarks for 1.5 MGD

	1.5 MGD								
Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Chemical dosage (ton/yr)	35	33	1	1	1	71	42	47	60
Total ThOD (ton/yr)	1,270	1,270	1,270	1,270	1,270	1,270	1,270	1,270	1,270
ThOD (BNR) (ton/yr)	1,066	1,029	1,050	1,048	1,045	1,138	1,101	997	1,448
Aeration (MWh/yr)	303	397	516	437	755	344	344	344	524
Pumping/RAS/Fluidization (BNR) (MWh/yr)	169	436	252	252	402	118	211	205	845
Total Energy (kWh/yr)	781	1,170	1,340	1,260	1,630	1,088	1,120	1,000	1,633
Aeration (kWh/kgThOD) (BNR)	0.28	0.39	0.49	0.42	0.72	0.30	0.44	0.35	0.36
Aeration (kWh/kg/kgN)	3.72	5.59	5.86	4.92	7.67	3.91	7.02	5.14	7.78
Pumping/Ras/Fluidization (kWh/kgThOD)	0.158	0.424	0.240	0.240	0.385	0.104	0.192	0.206	0.584
Total Energy/ThOD (kWh/kgThOD)	0.732	1.137	1.276	1.202	1.559	0.956	1.017	1.003	1.128
OpEx/ThOD ($/kgThOD) (BNR)	0.72	0.94	0.71	0.68	0.81	0.85	0.64	0.70	0.45
Total OpEx/gal ($/gal)	1.32	1.49	1.22	1.20	1.29	1.49	1.29	1.26	1.06
NPV/ThOD ($/kgThOD) (30-yr, 5%)	1.00	1.14	0.93	0.92	0.99	1.11	1.00	0.98	0.77
Solids Disposal OpEx ($1000/yr)	124.00	124.00	127.00	127.00	134.00	124.00	124.00	124.00	129.00
Chem P OpEx ($1000/yr)	120	114	3	3	4	163	96	109	138
Labour OpEx ($1000/yr)	993	1,170	962	954	995	1,220	1,030	960	776
Energy OpEx ($1000/yr)	78	117	134	126	163	109	112	100	163
Total OpEx ($1000/yr)	1,978	2,234	1,825	1,793	1,934	2,228	1,932	1,890	1588

Table A1.8 Energy Demand, Chemical Dosage, and OpEx Benchmarks for 4.6 MGD

	4.6 MGD								
Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Chemical dosage (ton/yr)	177	165	4	4	2	128	128	144	183
Total ThOD (ton/yr)	3,894	3,894	3,894	3,894	3,894	3,894	3,894	3,894	3,894
ThOD (BNR) (ton/yr)	3,627	3,153	3,599	3,424	3,599	3,501	3,383	3,124	4,352
Aeration (MWh/yr)	1,350	1,728	1,280	1,100	1,850	920	1,030	980	1,570
Pumping/RAS/Fluidization (BNR) (MWh/yr)	1,032	2,170	1,243	1,245	1,397	157	647	628	2,535
Total Energy (kWh/yr)	3,460	4,930	3,520	3,900	4,810	2,810	3,230	2,690	4,625
Aeration (kWh/kgThOD)	0.37	0.55	0.36	0.32	0.51	0.27	0.29	0.48	0.36
Aeration (kWh/kg/kgN)	5.41	7.93	4.74	4.03	6.13	3.44	4.60	7.26	7.60
Pumping/Ras/Fluidization (kWh/kgThOD)	0.285	0.688	0.345	0.364	0.388	0.045	0.191	0.201	0.582
Total Energy/ThOD (kWh/kgThOD)	0.954	1.564	0.978	1.139	1.336	0.803	0.955	0.861	1.063
OpEx/ThOD ($/kgThOD) (BNR)	0.47	0.63	0.40	0.40	0.46	0.50	0.47	0.53	0.47
Total OpEx/gal ($/gal)	0.90	0.93	0.80	0.85	0.93	0.86	0.84	0.79	0.77
NPV/ThOD ($/kgThOD) (30-yr, 5%)	0.70	0.72	0.64	0.69	0.74	0.66	0.66	0.63	0.55
Solids Disposal ($1000/yr)	166.00	162.00	163.00	163.00	209.00	162.00	161.00	161.00	164.00
Chem P OpEx ($1000/yr)	374	350	9	9	13	296	296	333	423
Labour OpEx ($1000/yr)	1,600	1,810	1,500	1,510	1,590	1,910	1,670	1,540	1,240
Energy OpEx ($1000/yr)	346	493	352	390	481	281	323	269	463
Total OpEx ($1000/yr)	4,120	4,300	3,661	3,919	4,270	3,973	3,851	3,645	3524

Table A1.9 Energy Demand, Chemical Dosage, and OpEx Benchmarks for 10 MGD

	10 MGD								
Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Chemical dosage (ton/yr)	326	331	9	4	9	279	279	313	427
Total ThOD (ton/yr)	8,466	8,466	8,466	8,466	8,466	8,466	8,466	8,466	8,466
ThOD (BNR) (ton/yr)	7,076	6,841	7,634	7,621	7,658	7,560	7,360	6,791	10,153
Aeration (MWh/yr)	2,760	3,373	3,350	3,210	4,870	2,130	2,290	2,180	3,670
Pumping/RAS/Fluidization (BNR) (MWh/yr)	1,505	4,338	2,360	2,698	2,700	340	1,449	1,360	5,914
Total Energy (kWh/yr)	6,280	9,760	7,980	8,440	10,400	7,660	6,540	7,260	10,534
Aeration (kWh/kgThOD)	0.39	0.49	0.44	0.42	0.64	0.28	0.44	0.32	0.36
Aeration (kWh/kg/kgN)	5.08	7.12	5.71	5.42	7.42	3.66	7.02	4.89	8.17
Pumping/Ras/Fluidization (kWh/kgThOD)	0.213	0.634	0.309	0.354	0.353	0.045	0.197	0.200	0.583
Total Energy/ThOD (kWh/kgThOD)	0.888	1.427	1.045	1.107	1.358	1.013	0.889	1.069	1.038
OpEx/ThOD ($/kgThOD) (BNR)	0.45	0.56	0.37	0.35	0.42	0.38	0.43	0.49	0.46
Total OpEx/gal ($/gal)	0.72	0.79	0.67	0.66	0.71	0.68	0.72	0.71	0.74
NPV/ThOD ($/kgThOD) (30-yr, 5%)	0.57	0.63	0.55	0.54	0.58	0.53	0.58	0.57	0.55
Solids Disposal ($1000/yr)	210.00	213.80	207.50	207.50	208.20	212.00	212.00	211.10	214.00
Chem P OpEx ($1000/yr)	752.00	761.00	20.50	28.10	20.50	643.00	643.00	725.00	983.00
Labour OpEx ($1000/yr)	2,220	2,610	2,160	2,180	2,230	2,690	2,510	2,350	1,870
Energy OpEx ($1000/yr)	628	976	798	844	1,040	766	654	726	1,053
Total OpEx ($1000/yr)	7,180	7,892	6,679	6,552	7,094	6,789	7,179	7,051	7396

Table A1.10 Energy Demand, Chemical Dosage, and OpEx Benchmarks for 20 MGD

	20 MGD								
Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Chemical dosage (ton/yr)	707	661	18	18	18	558	558	626	790
Total ThOD (ton/yr)	16,932	16,932	16,932	16,932	16,932	16,932	16,932	16,932	16,932
ThOD (BNR) (ton/yr)	15,070	13,715	15,301	15,277	15,350	15,121	14,720	13,557	18,841
Aeration (MWh/yr)	5,120	6,609	5,900	5,220	8,530	4,250	4,580	4,360	6,810
Pumping/RAS/Fluidization (BNR) (MWh/yr)	4,144	8,652	5,404	5,404	6,074	273	2,897	2,720	10,980
Total Energy (kWh/yr)	13,700	18,800	14,600	15,600	19,600	11,230	12,800	10,800	19,060
Aeration (kWh/kgThOD)	0.34	0.48	0.39	0.34	0.56	0.30	0.44	0.32	0.36
Aeration (kWh/kg/kgN)	4.72	6.97	5.02	4.40	6.50	3.83	7.01	4.89	7.58
Pumping/Ras/Fluidization (kWh/kgThOD)	0.275	0.631	0.353	0.354	0.396	0.018	0.197	0.201	0.583
Total Energy/ThOD (kWh/kgThOD)	0.909	1.371	0.954	1.021	1.277	0.743	0.870	0.797	1.012
OpEx/ThOD ($/kgThOD) (BNR)	0.40	0.46	0.30	0.29	0.36	0.33	0.39	0.45	0.44
Total OpEx/gal ($/gal)	0.67	0.68	0.57	0.56	0.62	0.60	0.65	0.64	0.62
NPV/ThOD ($/kgThOD) (30-yr, 5%)	0.54	0.54	0.47	0.46	0.51	0.48	0.52	0.52	0.45
Solids Disposal ($1000/yr)	347.00	418.50	289.60	289.60	289.60	415.00	414.90	412.50	312.00
Chem P OpEx ($1000/yr)	1630.00	1520.00	41.00	41.00	41.00	1290.00	1290.00	1450.00	1820.00
Labour OpEx ($1000/yr)	3,700	3,940	3,230	3,300	3,390	4,100	3,980	3,750	2,780
Energy OpEx ($1000/yr)	1,370	1,880	1,460	1,560	1,960	1,123	1,280	1,080	1,906
Total OpEx ($1000/yr)	13,489	13,535	11,327	11,243	12,342	11,931	12,912	12,855	12,359

Table A1.11 Energy Demand, Chemical Dosage, and OpEx Benchmarks for 30 MGD

	30 MGD								
Parameters	MLE	4BDP	A2O	UCT	5BDP	BAF	IFAS	MBBR	CFBBR
Chemical dosage (ton/yr)	1,060	992	27	27	27	837	837	940	1,210
Total ThOD (ton/yr)	25,398	25,398	25,398	25,398	25,398	25,398	25,398	25,398	25,398
ThOD (BNR) (ton/yr)	22,555	20,557	22,936	22,899	23,009	22,707	22,097	20,349	28,960
Aeration (MWh/yr)	7,750	9,845	8,990	9,370	13,000	6,400	6,870	6,530	10,500
Pumping/RAS/Fluidization (BNR) (MWh/yr)	6,177	12,990	7,090	8,100	8,100	1,020	4,345	4,080	16,900
Total Energy (kWh/yr)	18,900	27,800	21,500	24,800	29,400	15,740	18,400	15,500	29,300
Aeration (kWh/kgThOD)	0.34	0.48	0.39	0.41	0.57	0.27	0.44	0.32	0.36
Aeration (kWh/kg/kgN)	4.76	6.93	5.10	5.27	6.60	3.52	7.01	4.88	7.80
Pumping/Ras/Fluidization (kWh/kgThOD)	0.274	0.632	0.309	0.354	0.352	0.045	0.197	0.201	0.584
Total Energy/ThOD (kWh/kgThOD)	0.838	1.352	0.937	1.083	1.278	0.693	0.833	0.762	1.012
OpEx/ThOD ($/kgThOD) (BNR)	0.38	0.46	0.28	0.29	0.35	0.31	0.37	0.43	0.43
Total OpEx/gal ($/gal)	0.64	0.66	0.53	0.54	0.59	0.56	0.61	0.61	0.61
NPV/ThOD ($/kgThOD) (30-yr, 5%)	0.51	0.53	0.45	0.45	0.49	0.45	0.49	0.49	0.45
Solids Disposal ($1000/yr)	550.00	486.00	360.00	360.00	360.00	486.00	481.00	478.00	368.00
Chem P OpEx ($1000/yr)	2440.00	2280.00	61.40	61.40	61.40	1930.00	1930.00	2180.00	2790.00
Labour OpEx ($1000/yr)	4,900	5,160	4,240	4,340	4,470	5,360	5,340	5,030	3,770
Energy OpEx ($1000/yr)	1,890	2,780	2,150	2,480	2,940	1,574	1,840	1,550	2,930
Total OpEx ($1000/yr)	19,257	19,737	16,047	16,238	17,631	16,911	18,162	18,165	18323

CapEx and OpEx Charts

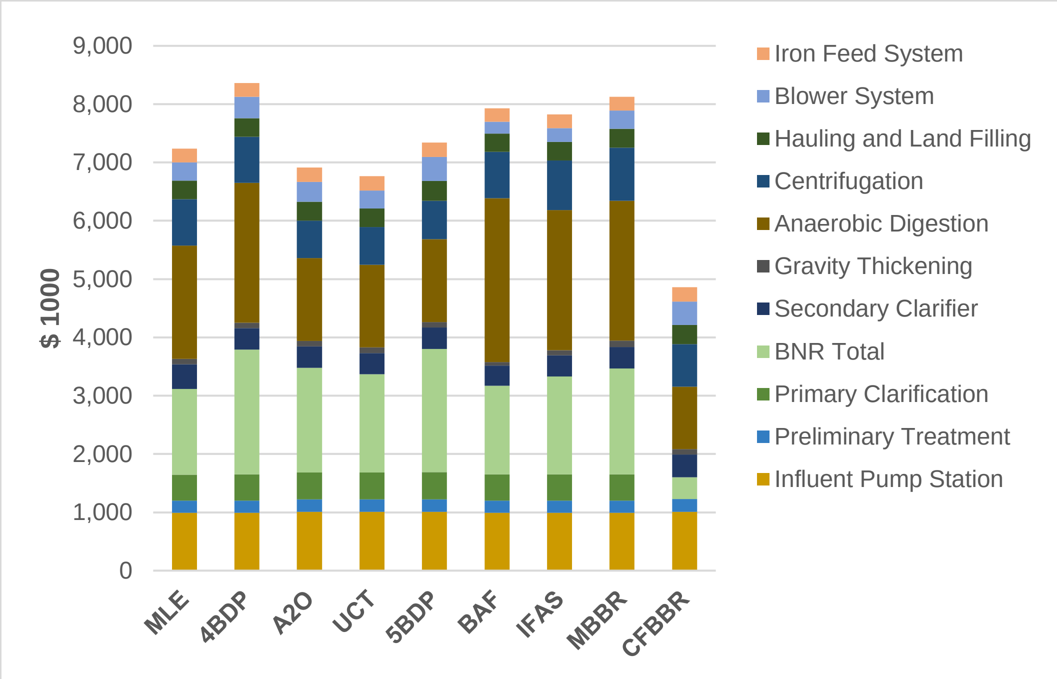

Figure A1.1 CapEx breakdown at 1.5 MGD

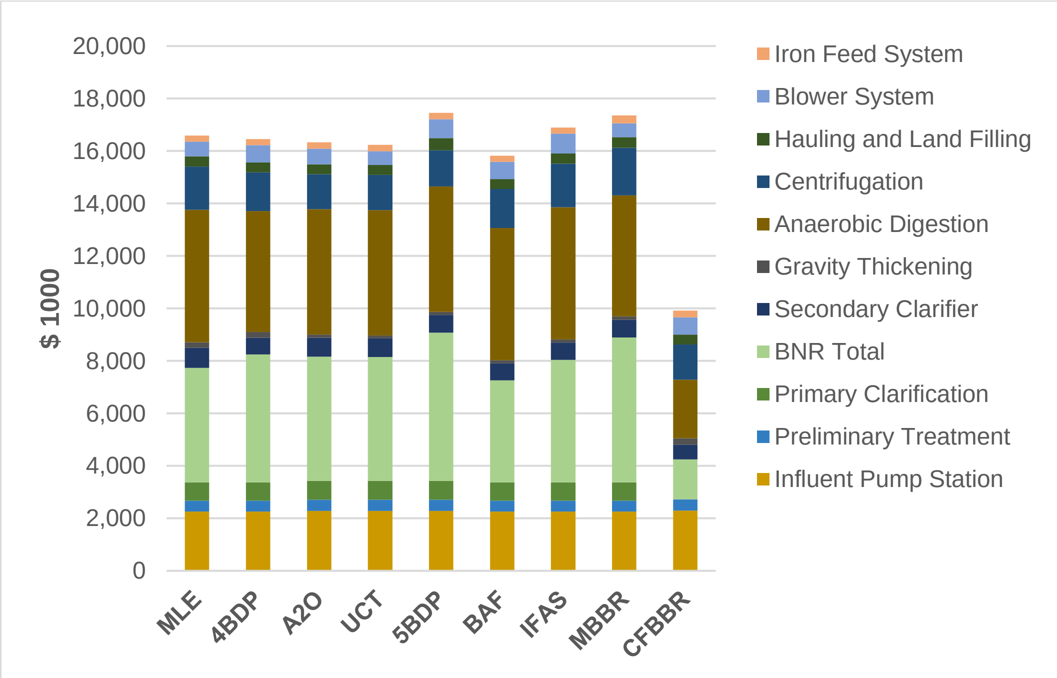

Figure A1.2 CapEx breakdown at 4.6 MGD

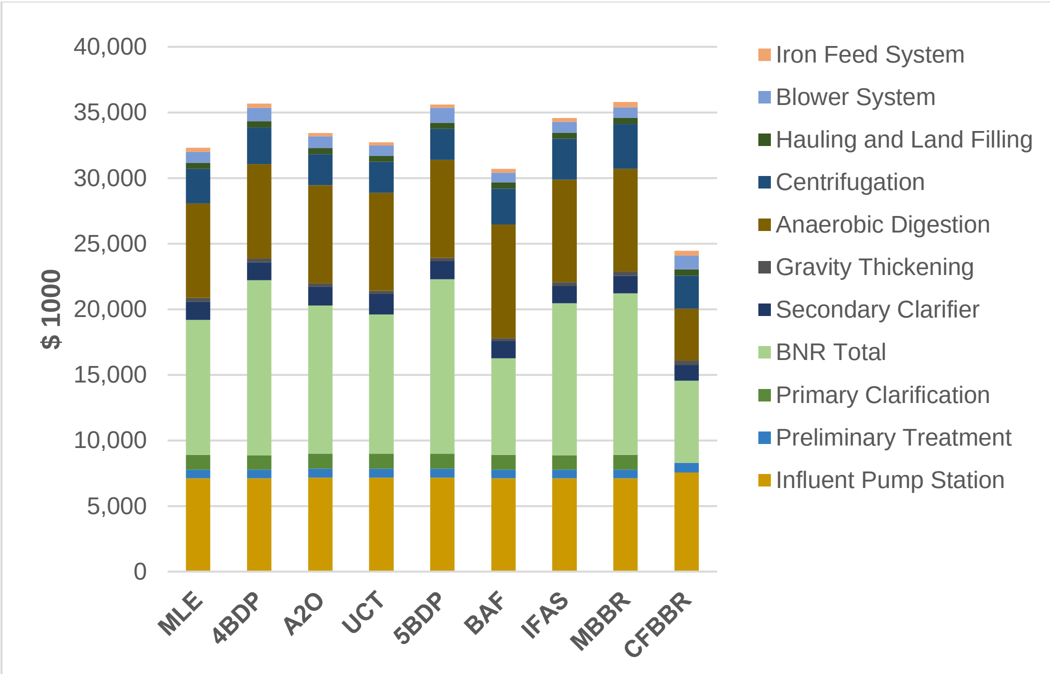

Figure A1.3 CapEx breakdown at 10 MGD

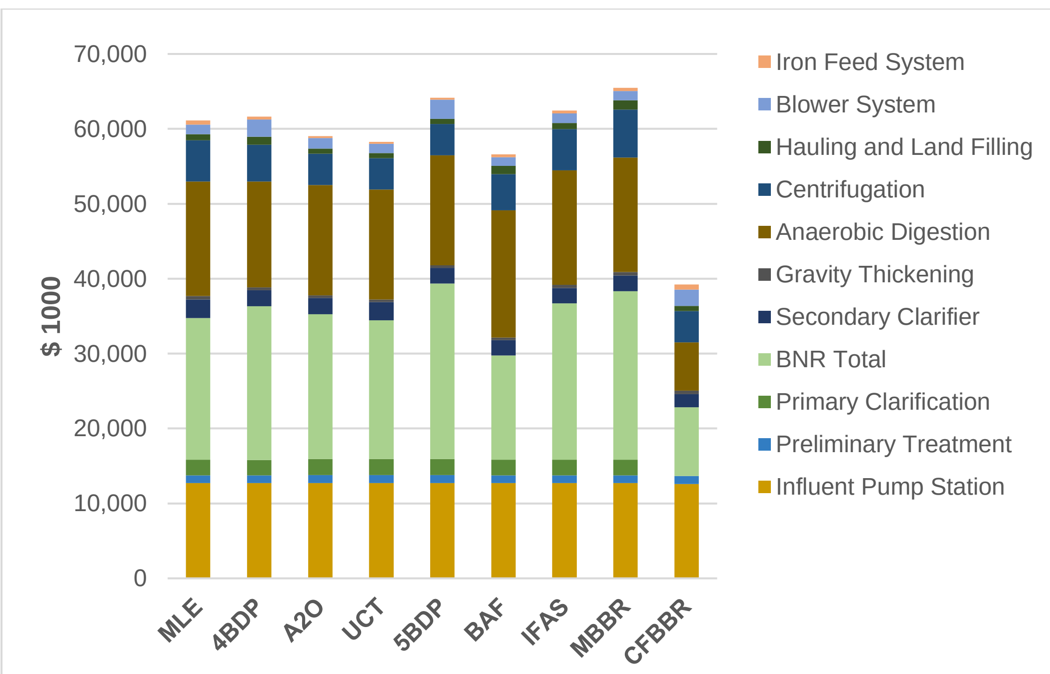

Figure A1.4 CapEx breakdown at 20 MGD

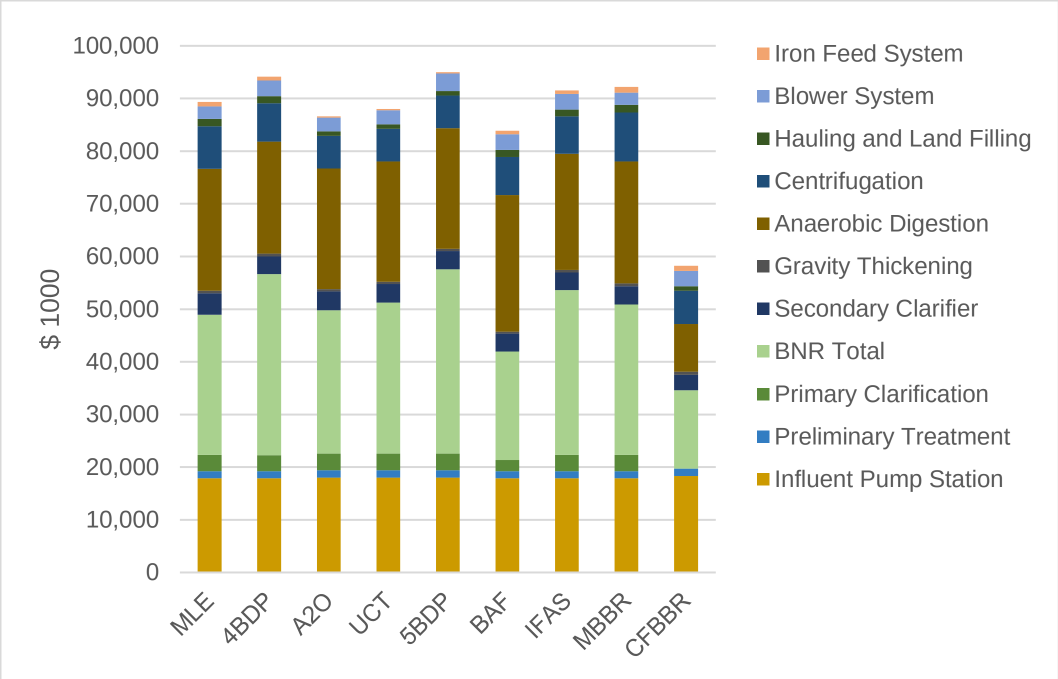

Figure A1.5 CapEx breakdown at 30 MGD

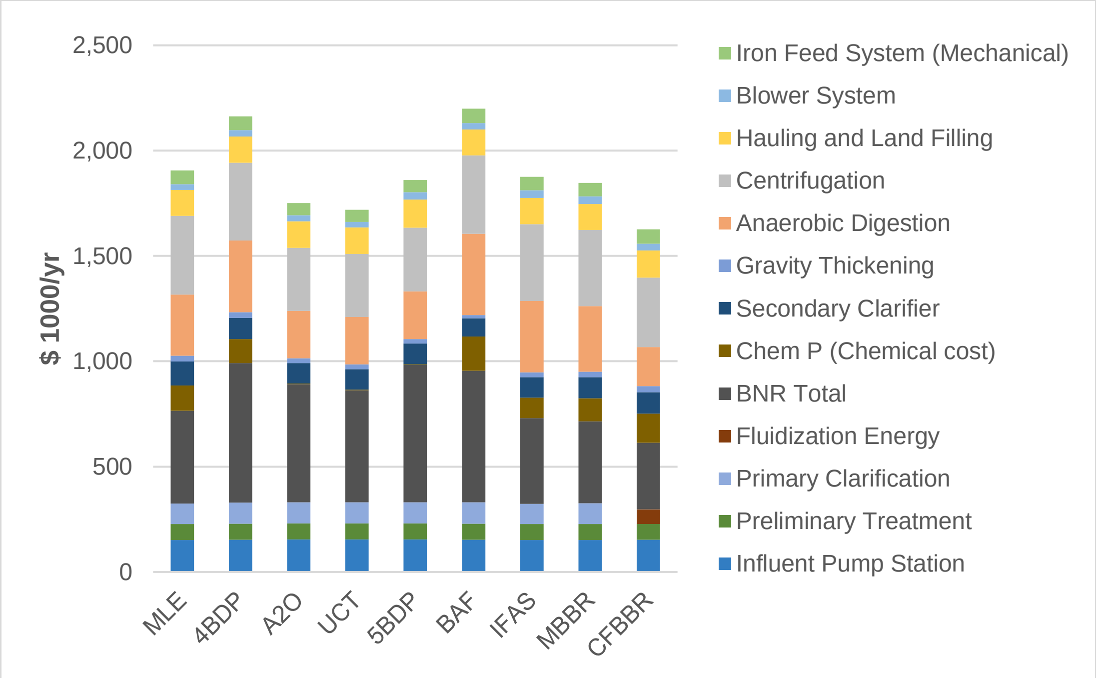

Figure A1.6 OpEx breakdown at 1.5 MGD

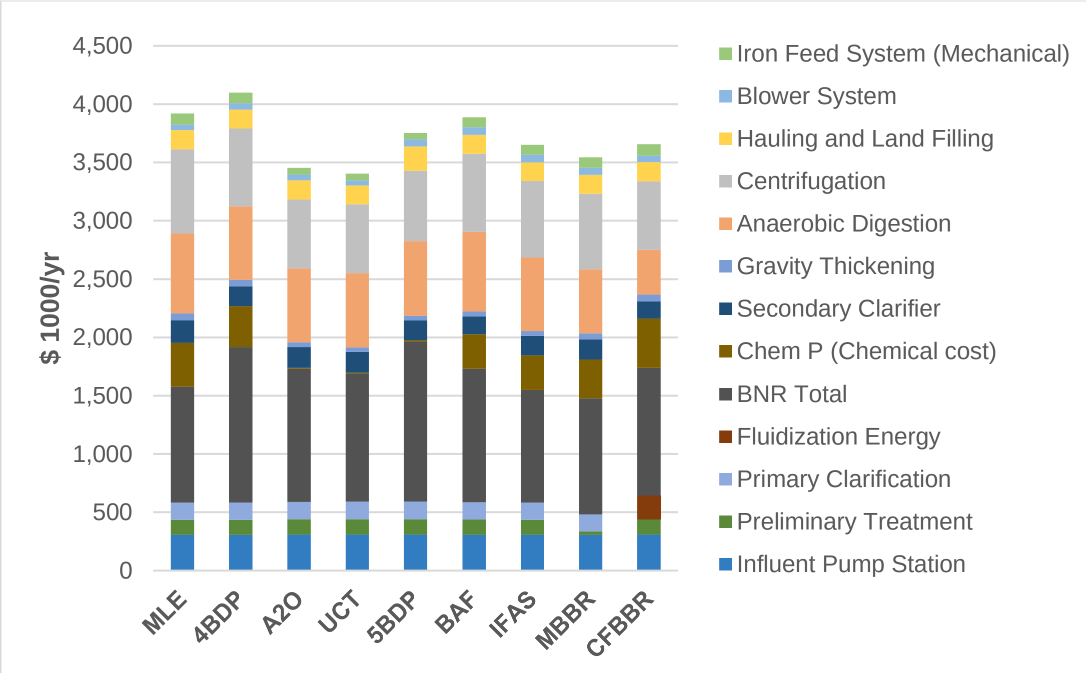

Figure A1.7 OpEx breakdown at 4.6 MGD

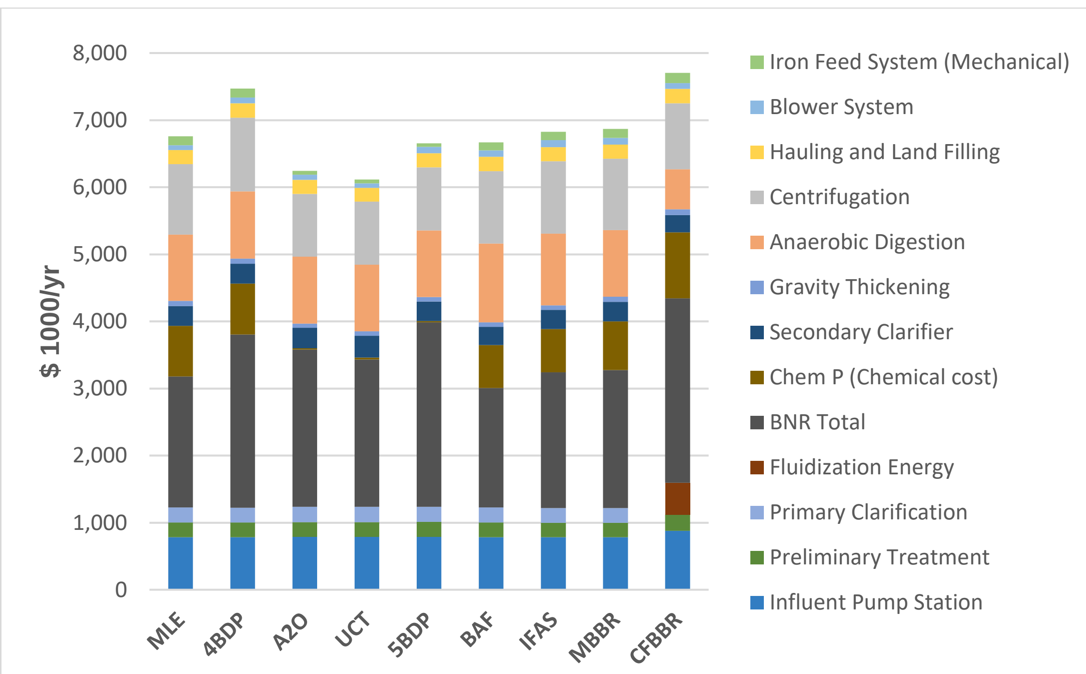

Figure A1.8 OpEx breakdown at 10 MGD

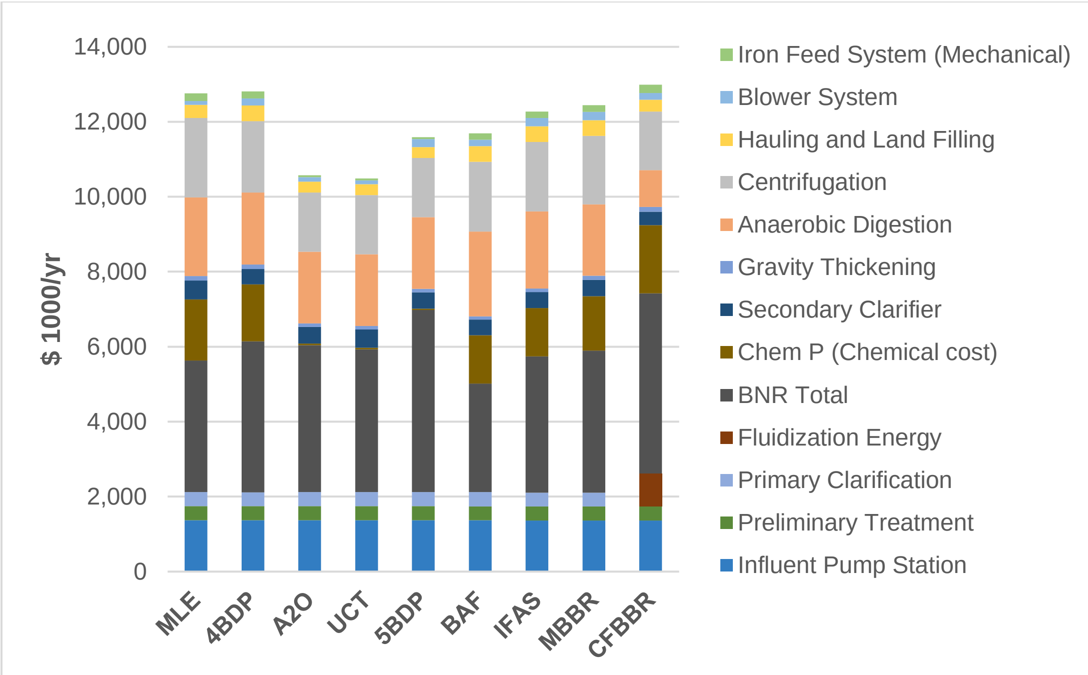

Figure A1.9 OpEx breakdown at 20 MGD

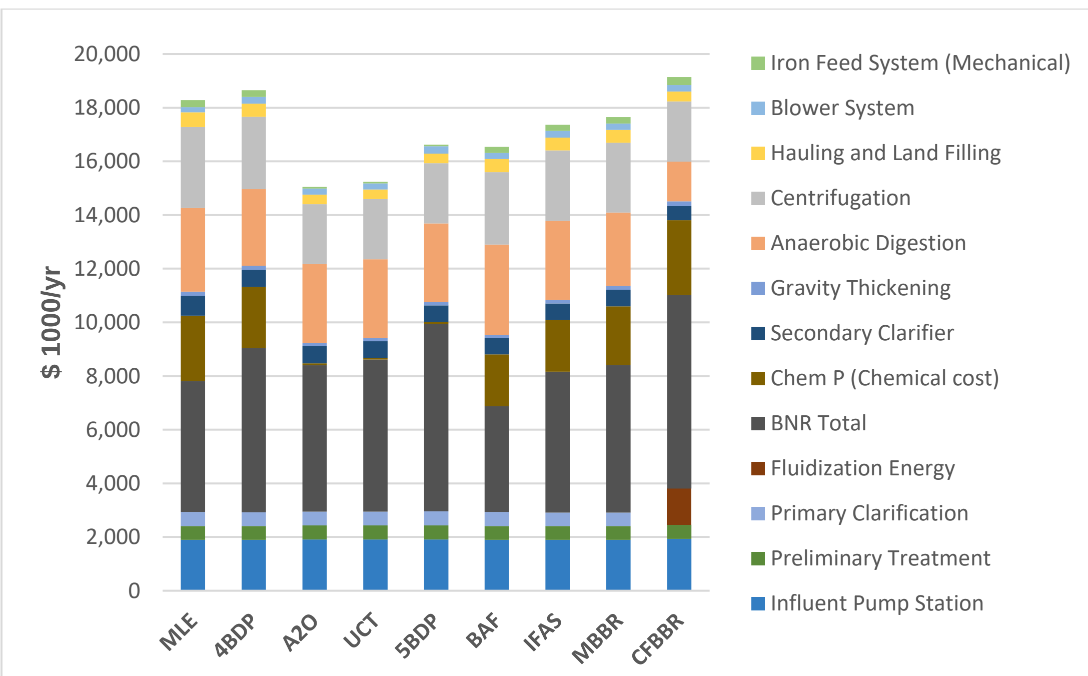

Figure A1.10 OpEx breakdown at 30 MGD

Net Present Value Charts

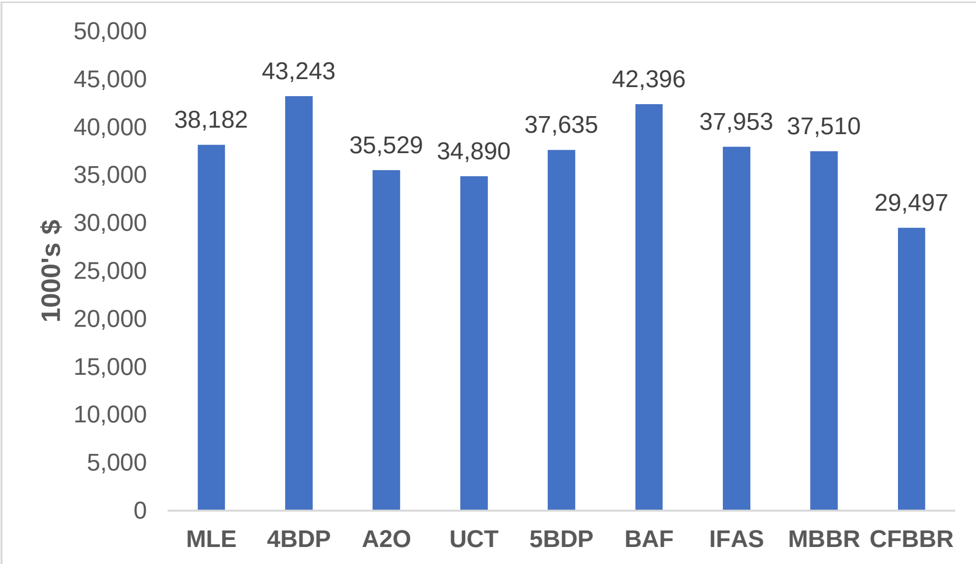

Figure A1.11 Net present value at 1.5 MGD

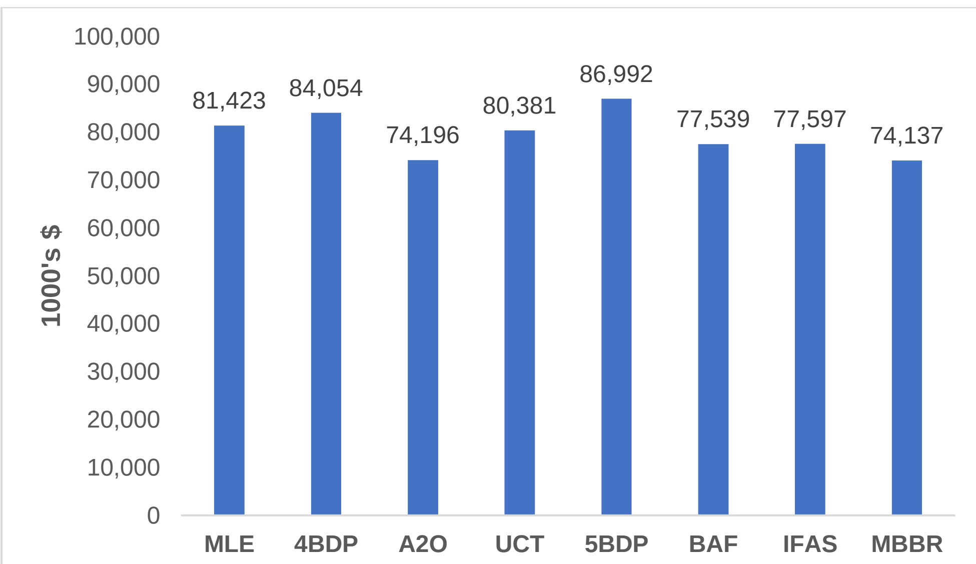

Figure A1.12 Net present value at 4.6 MGD

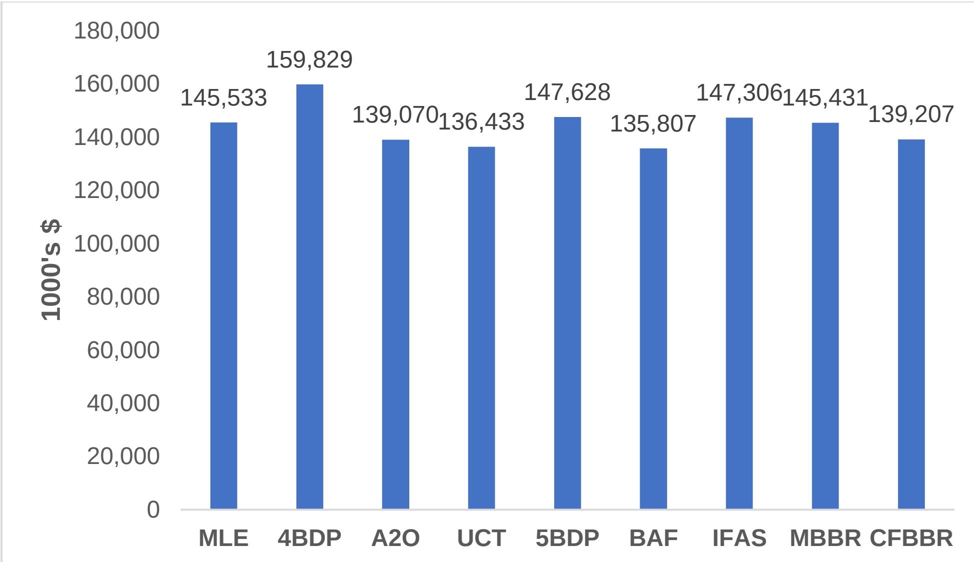

Figure A1.13 Net present value at 10 MGD

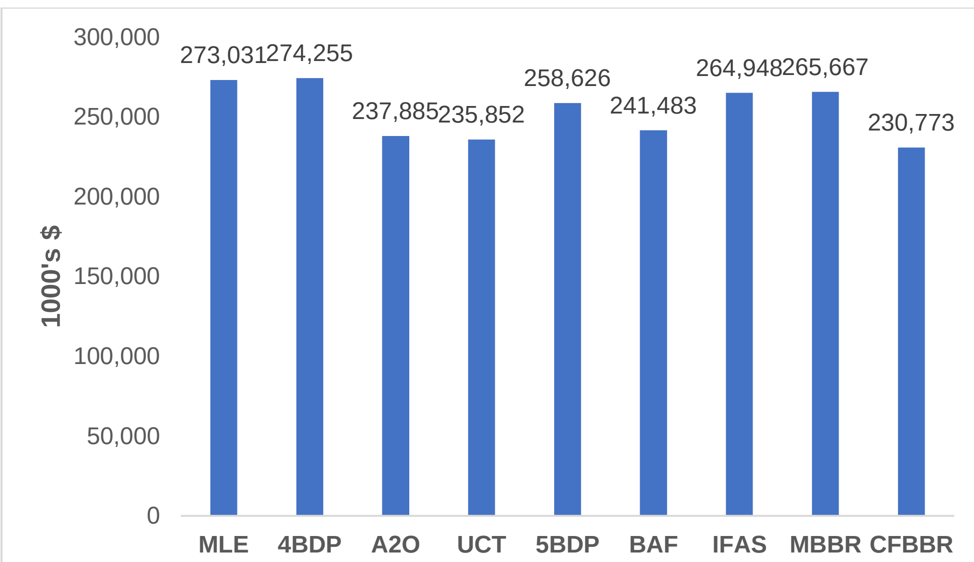

Figure A1.14 Net present value at 20MGD

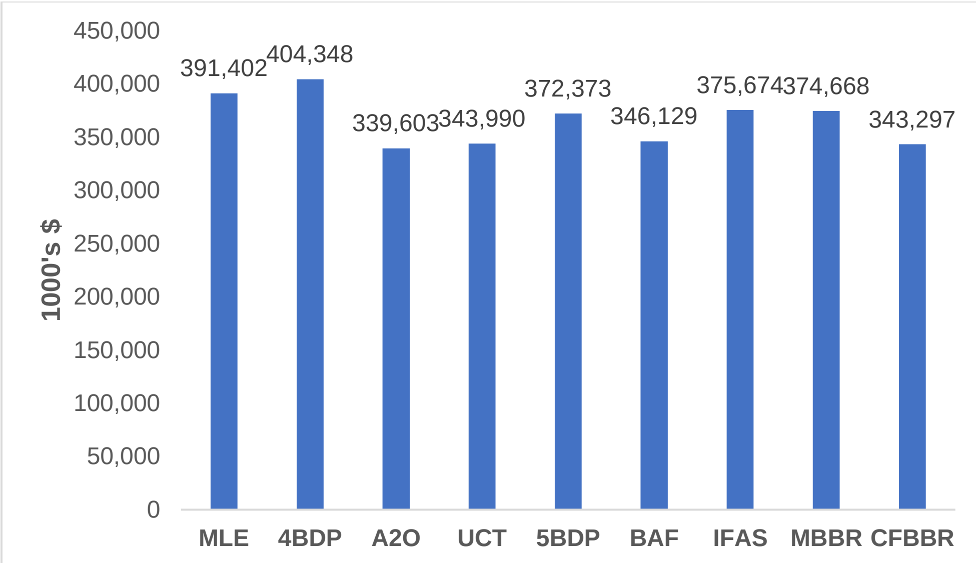

Figure A1.15 Net present value at 30 MGD

Table A1.12 CapEx and OpEx Units Costs

Unit Costs **Description**	**Value**	**Units**
Building Cost	110	$/sqft
Excavation	8	$/cuyd
Wall Concrete	650	$/cuyd
Slab Concrete	350	$/cuyd
Crane Rental	250	$/hr
Canopy Roof	20	$/sqft
Electricity	0.1	$/kWh
Hand Rail	75	$/ft
Land Costs	20000	$/acre
Construction Labor Rate	40	$/hr
Operator Labor Rate	51.5	$/hr
Administration Labor Rate	51.5	$/hr
Laboratory Labor Rate	51.5	$/hr
Hydrated Lime-[Ca(OH)2]	0.18	$/lb
Al2(SO4)3*14H2O	0.27	$/lb
Ferric Chloride	0.36	$/lb
Polymer	1.3	$/lb
Citric Acid 50%	0.52	$/lb
NaOCl 14%	9.76	$/cuft
Interest Rate	8	%
Construction Period	3	years
Operating Life of Plant	40	years
Engineering Design Fee	15	%
Miscellaneous	5	%
Administration/Legal	2	%
Inspection	2	%
Contingency	10	%
Technical	2	%
Profit and Overhead	15	%

Appendix 2 (Chapter 5)

Biomass Yield Cumulative Graphs

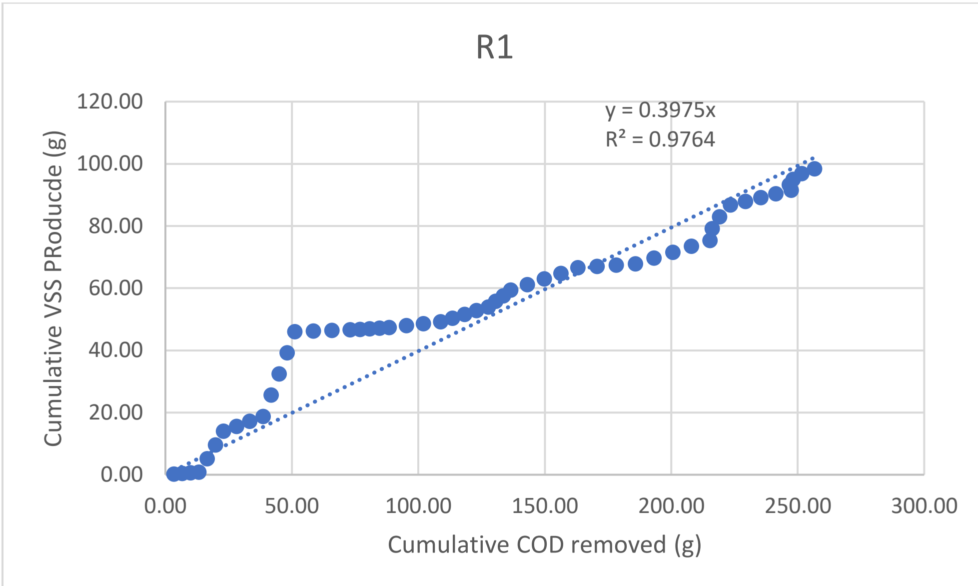

Figure A2.1 Yield Chart of R1

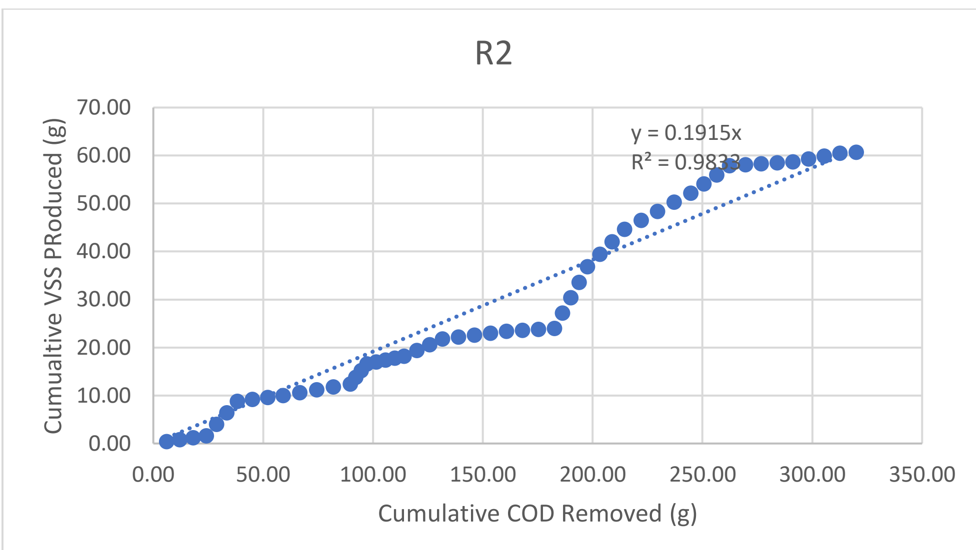

Figure A2.2 Yield Chart of R2

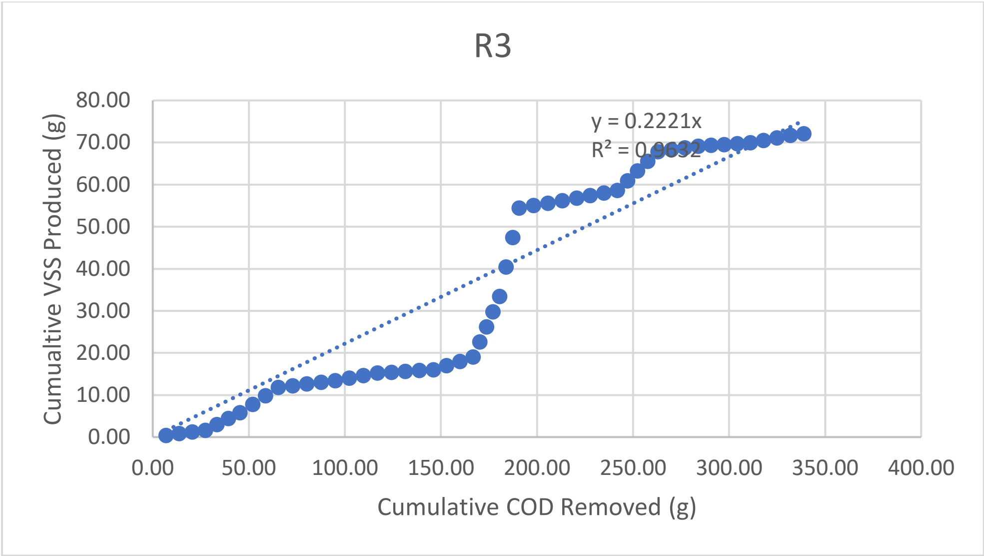

Figure A2.3 Yield Chart of R3

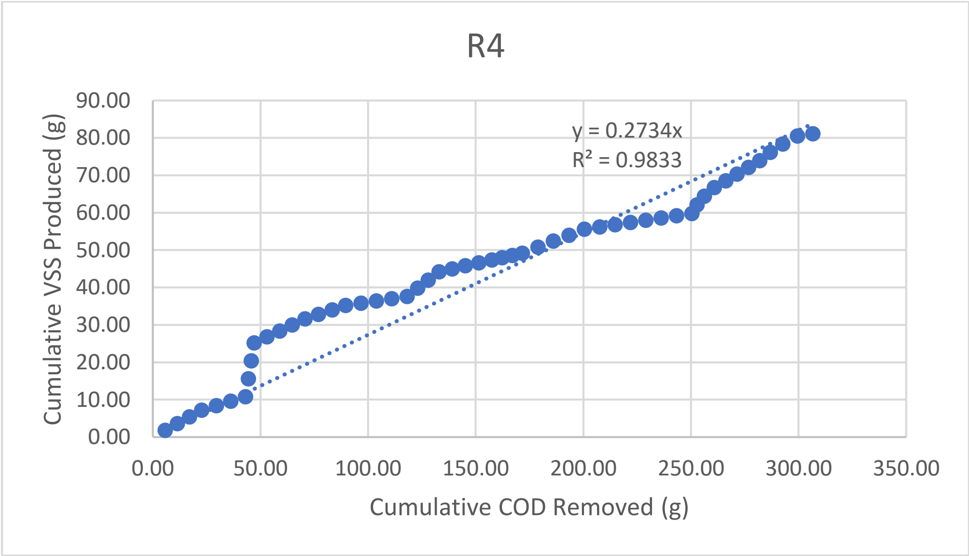

Figure A2.4 Yield Chart of R4

Appendix 2

Specific Nitrification Rate Tests

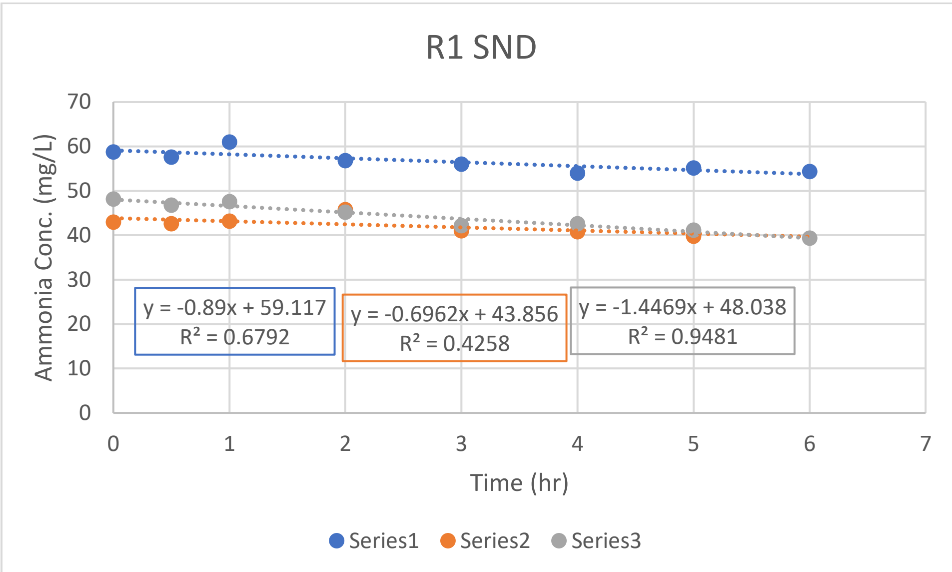

Figure A2.5 SND of R1

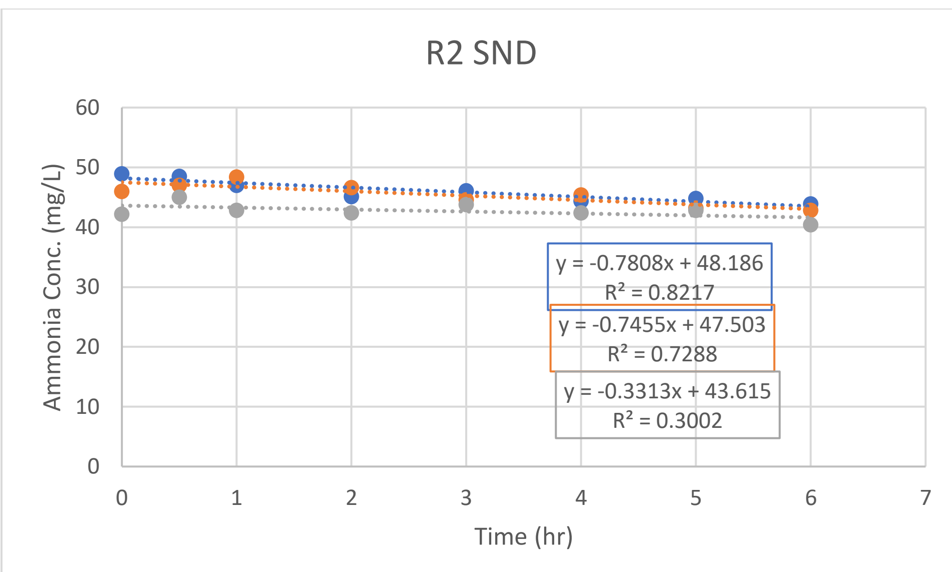

Figure A2.6 SND of R2

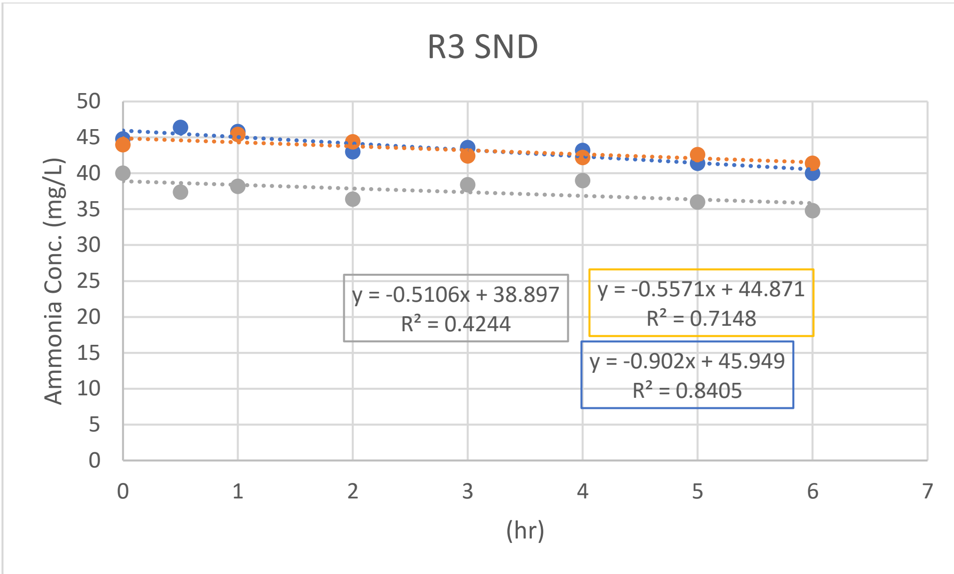

Figure A2.7 SND of R3

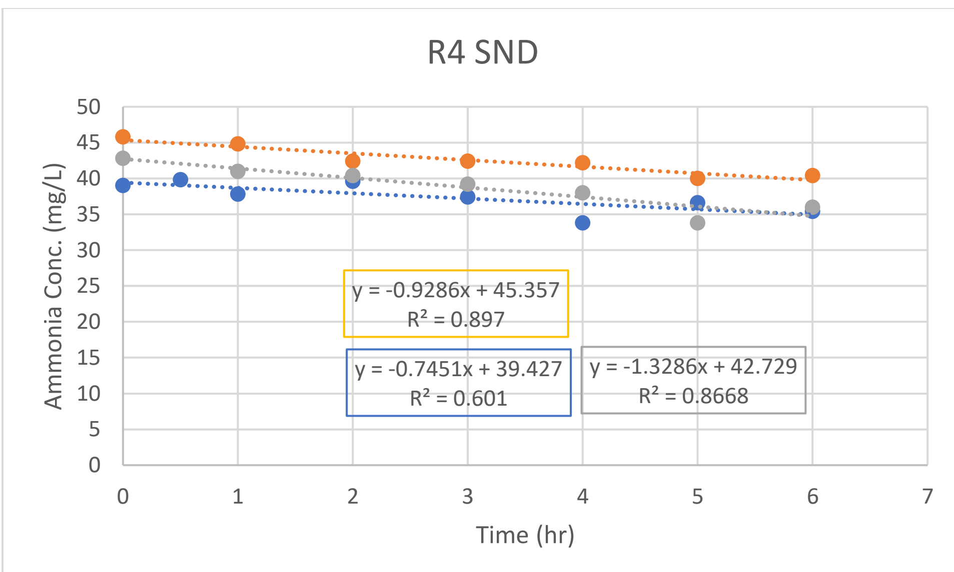

Figure A2.8 SND of R4

Curriculum Vitae

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Education

Ph.D. Candidate, Chemical and Biochemical Engineering **09/2016 – 2021**
Western University, London, Ontario & Tianjin University, Tianjin, China

- Dual-Degree PhD Program between Western University and Tianjin University

Master of Engineering, Chemical and Biochemical Engineering **09/2015 – 08/2016**
Western University, London, Ontario

Bachelor of Engineering Science, Green Process Engineering **09/2011 – 06/2015**
Western University, London, Ontario

Publications

- **Nelson, M.J.**, Nakhla, G., Zhu, J. (2017) Fluidized-Bed Bioreactor Applications for Biological Wastewater Treatment: A Review of Research and Developments. Engineering. 3: 330-342.
- Liu, A., **Nelson, M.J.**, Wang, X., Li, H., He, X., Zhao, Z., Zhong, H., Nakhla, G., Zhu, J. (2019) Decentralized Wastewater Treatment in an Urban Setting: A Pilot Study of the Circulating Fluidized Bed Bioreactor Treating Septic Tank Effluent. Environmental Sciences.
- **Nelson, M.J.**, Nakhla, G., Zhu, J. (2021) The Circulating Fluidized Bed Bioreactor as a Biological Nutrient Removal Process for Municipal Wastewater Treatment: Process Modelling and Costing Analysis. Journal of Environmental Management. Accepted August 2021.

Awards & Scholarships

- Mitacs Globalink Research Award: Awarded May 2017; Received August 2017; Value $5000
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Work Experience

Graduate Teaching Assistant **09/2016 – 04/2021 (Intermittent)**
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