MOVING BED BIOFILM REACTOR FOR ONSITE TREATMENT OF WINERY WASTEWATER

Final Technical Report

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Executive Summary

The design outlined in this report is tailored to address the two overriding challenges for wastewater treatment at small wineries: highly variable hydraulic and organic loading rates. We have found that the Moving Bed Bioreactor (MBBR) technology shows promise as an economically feasible and mechanically reliable option that will help wineries to comply with state and federal regulations, reduce costs associated with waste management, and reduce the environmental footprint of their product. To test the practicality of using an MBBR in treating winery wastewater the team constructed a bench scale prototype featuring a 9" EPDM diffuser and cylindrical tank geometry. Our team then used the prototype to run a series of experiments focused on chemical oxygen demand (COD) analysis and optimization of the aeration system. The results showed an increasing, but variable trend in COD removal, with a maximum removal of 82.3%. To confirm the COD analysis our group also analyzed the microbial respiration rate in a series of dissolved oxygen (DO) consumption tests. The results of these tests were modeled and used to determine the aeration coefficient of wastewater to clean water. From our prototype analysis and modeled values, we determined that one 435 gallon MBBR would be sufficient to treat the maximum hydraulic and organic load of 1000 gal/day and 5,000 mg/L BOD respectively.

Introduction

Pacific Northwest wineries are growing rapidly, with over 19,000 acres of vineyards planted in Oregon alone [2]. The process from grape to bottle is dominated by seasonal periods where the winery become highly productive, then followed periods of relative inactivity. The peak operation period, known as "crush" typically occurs during the late summer to early fall. During this time the wineries are processing the grapes into juice in preparation for the fermentation.

Small wineries (8,000 cases per year) can typically produce anywhere from 500 – 1000 gallons per day of wastewater during this time [4]. The wastewater produced during the crush period and at other processing times such as during barrel washes can contain very high loads of biochemical oxygen demand (BOD). In the past, much of this waste stream went untreated; however due to increasing number of wineries, many municipalities are beginning to require treatment to maintain the health of local water bodies. Wineries located off the municipal grid may be required to treat wastewater onsite for discharge to drain fields or pay fees for collecting, hauling, and treating the waste. As expenses and complying with regulatory requirements are vital to the management of a business, many of these wineries are now considering alternatives to their current wastewater management operations.

Wastewater Treatment Challenges

Wineries produce an estimated 1000 gallons of wastewater per 264 gallons of wine [1]. The wastewater produced can have highly variable chemical oxygen demand (COD), biochemical oxygen demand (BOD) and total suspended solids (TSS).

The sources of the BOD and TSS come from compounds composed of alcohols, acids, sugars, tannins and lignin [2]. COD sources come from any chemicals entering the waste stream that can be oxidized. Of the total COD, ~70% can be assumed to be biological [3]. Winery wastewater BOD concentrations can be as high as 20,000 mg/L and concentrations of TSS can range from 300 to 30,000 mg/L [1].

High concentrations of BOD and TSS can be detrimental to aquatic life, disrupting ecological processes, creating eutrophic conditions and upsetting the balance of nutrient systems. Because of this, environmental regulations may impact the ways in which wineries can dispose of their wastewaters, and in most circumstances, treatment of the wastewater is needed to reduce BOD and TSS concentrations prior to disposal into the municipal waste system. The costs that wineries incur in wastewater disposal can have a significant impact on their business. Disposal costs can be significantly reduced if a primary wastewater treatment system is implemented to reduce contaminant loads.

Parameter	Units	Winery Waste Characteristics Range During Vintage			Winery Waste Characteristics Range During Non-Vintage		
		Minimum	Average	Maximum	Minimum	Average	Maximum
Flow	% of Total	-	63%	-	-	37%	-
Load Duration	Days	-	75	-	-	290	-
Wastewater Produced	gallons/ton of grapes	1100					
Chlorides	mg/l	10	20	60	20	600	5000
Total Solids	mg/l	146	756	1,250	347	600	1,127
Suspended Solids	mg/l	5	300	660	5	260	720
Volatile Suspended Solids	%	-	38%	-	-	49%	-
Settleable Solids	ml/l	1	4	8	0	1	6
COD	mg/l	386	810	1,412	285	825	2,880
BOD	mg/l	131	1,560	7,200	128	1,440	4,200
pH	s.u.	5.0	-	7.1	5.0	-	10.0
Total Kjeldahl Nitrogen	mg/l	1.0	2.4	5.7	1.2	4.7	12.4
Ammonia as N	mg/l as N	1.5	1.8	4.8	0.8	1.0	3.3
Nitrates	mg/l as N	0.9	1.5	2.3	0.1	0.3	0.5
Total Phosphorous	mg/l as P	0.5	1.2	2.5	0.3	1.0	1.3

FIGURE 1 NAPA VALLEY SANITATION DISTRICT WINERY WASTEWATER POLLUTANT CHARACTERISTICS

The above parameters in Figure 1, from Winery Waste Management Napa Sanitation District, were used as the base for our design values [8]. Figure 2 below, shows the average flow rate by month expected by wineries [8].

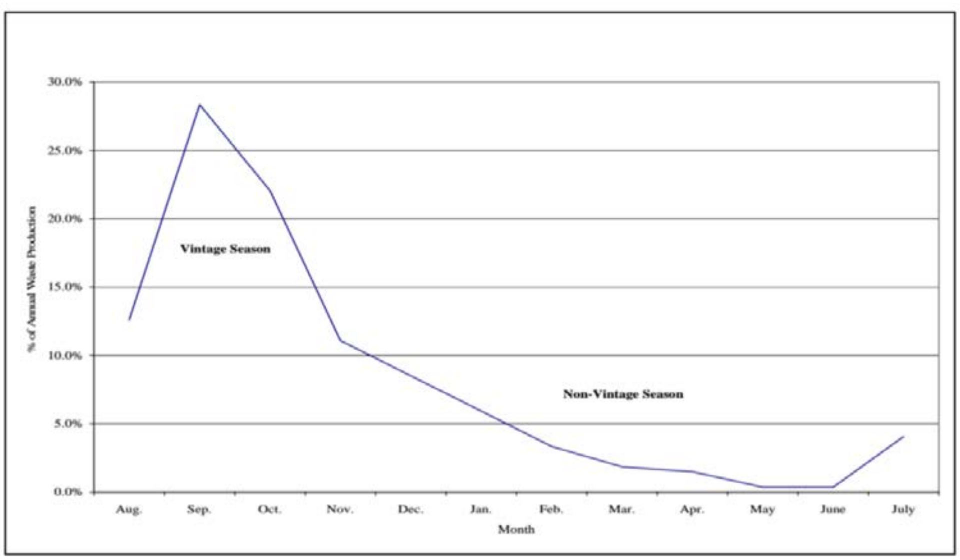

FIGURE 2 ANNUAL THEORETICAL WINERY WASTE PRODUCTION PERCENTAGE BY MONTH

Wastewater Treatment Solutions

Bioreactor systems for winery wastewater are designed for a specific purpose: to degrade pollutants in winery effluent by removing total suspended solids and biochemical oxygen demand from the process water. The process is most efficiently done by the microbial metabolic oxidation of *dissolved* organic material in the effluent.

Problem Statement

Eco Energetics Design Group will design a primary treatment system for high strength process wastewater from a mid-size vineyard. Vineyard operators require a system capable of dealing with seasonal variations in flow, high biological oxygen demand (BOD), and high total suspended solids (TSS) concentrations. The system will handle peak flow rates of 1000 gpd, maximum concentrations of TSS of 700 mg/L and a BOD load of 7000 mg/L during the vintage season. The system will reduce the BOD to 500 mg/L or less and TSS to 100 mg/L or less prior to entering the AdvanTex AX100 polishing system. These capabilities are only achievable post clarification and under normal pH ranges of 6-8, and in the absence of additional inorganic/organic chemicals such as cleaning products. The system will be designed to minimize maintenance and operational cost while maximizing effluent quality. Constraints for our design include the legal regulations, material availability, start-up time, prototype testability, economic analysis, and treatment efficiency. We selected an aerobic system due to their relatively short start up time and ability to deal with shock loadings. Both attached and suspended growth systems were considered, but ultimately a hybrid system was chosen that offers the benefits of both.

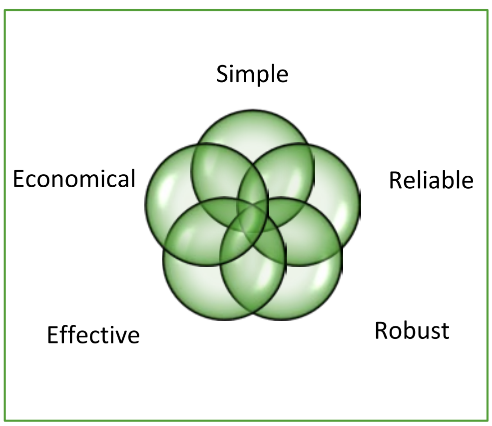

FIGURE 3 MBBR TREATMENT

Our team selected Moving a Bed Bioreactor (MBBR) as the primary treatment method. Media packed beds have been shown to achieve from 88-96% COD removal [3]. This puts the required removal efficiencies of 92% well within reach, provided there is adequate aeration and nutrient supplementation.

Per the Operational Guidelines proposed by Wine Australia for winery effluent management and reuse, MBBR treatment systems have the characteristics outlined in Figure 3 that make them ideal for our purposes [2].

The primary goal is to reduce BOD_5 to a level suitable for secondary polishing treatment with Orenco Systems patented AdvanTex textile media. A secondary goal is decreasing total suspended solids (TSS), this will be achieved in pre-and post- treatment clarification basins. This post clarification will be required to meet effluent quality requirements due to the increase of TSS in MBBR effluent which is proportional to the organic loading.

Background

Regulatory Requirements

Once the winery effluent is discharged from the treatment system it must be disposed of in a manner that follows both state and federal regulations. The methods of disposal listed below assume that the winery is not on a municipal sewer system:

- Storage of treated effluent onsite with periodic removal to the closest municipal wastewater facility.
- Discharge of treated effluent into a septic drain field.
- Storage for irrigation use.

The effluent discharges from a typical winery are classified as *high strength wastewater*, because the BOD exceeds 320 mg/L [4]. The unusually high BOD of winery wastewater (~7000 mg/L during crush) represents a significant cost to winery operations as municipal systems have charges that are assessed on BOD in excess of the first 300 mg/L, per 1,000 gallons of effluent sent to municipal treatment. It is recommended that effluent is treated and disposed of on-site to avoid these costs.

Disposal Regulations

- Process wastewater must be collected and adequately screened before it is land-applied on the property where it is generated.
- Sewage must be kept separate from process wastewater.
- The wastewater management plan must include information on agronomic land application practices, pond or evaporative pond storage design, treatment methods, and management of solids.
- Disposal and storage must be protective of groundwater.

The Oregon Department of Environmental Quality (DEQ) requires the 1400A permit for wineries that do not exceed 25,000 gallons per day [9]. The 1400A permit is required for wineries and seasonal fresh-pack food operations that do not significantly alter the final product for market. Wastewater generated from this type of food preparation is only disposed of by land application for beneficial reuse. No discharge can flow into surface water bodies.

Sludge Disposal and Land-Application Regulations

The primary regulatory concern for the MBBR is sludge disposal. There are four main ways to dispose of sludge: Land-spreading, other beneficial uses (such as composting), co-disposal with sewage sludge, and landfilling. A Toxicity Characteristic Leaching Procedure (TCLP) must be performed *prior to* sending the sludge to a landfill.

The current design recommends incorporating land-application sludge disposal for fertilization. Land application of sludge is currently regulated under the Resource Conservation and Recovery Act as well as state and local agencies [4]. Bio-solids are regulated under the DEQ's water quality program, specifically through a National Pollutant Discharge Elimination System (NPDES) or Water Pollution Control Facility (WPCF) permit.

Permitting and site approval requirements for bio-solids application to cropland depends on the following bio-solid characteristics: trace element concentrations of heavy metals and pathogens in the bio-solids relative to USEPA standard and whether the bio-solids meet USEPA Class A or Class B pathogen reduction requirements. Site approvals are *always* required for Class B bio-solids application. Site approvals *may* be required for Class A bio-solids application if trace element concentrations of heavy metals and pathogens in bio-solids exceed Exceptional Quality (EQ) limits, or if bio-solids are not stabilized sufficiently via digestion or composting. Treatment facilities producing Class A bio-solids are required to demonstrate that their treatment process meets time and temperature standards. They also are required to monitor bio-solids products for the presence of human pathogen indicator organisms. Specifically, a winery waste-stream is not likely to be either pathogenic or contain gross amounts of heavy metals. Simple water quality tests can determine eligibility for land application of sludge at an individual winery.

Technology Selection

Design Matrix

A weighted decision matrix (WDM) is a simple tool that can be very useful in making complex decisions, especially when many alternatives and criteria of varying importance are to be considered. We designed a comparative matrix which scored each technology based on important design characteristics (Figure 18). Each individual characteristic was assigned one of the following categories: economic, treatment, design & implementation, operation & maintenance, and 'other'. Each category was assigned a weight out of 100 based on importance. This weight was divided amongst the characteristics within that category.

The 'Ability to Handle Variable Flows and Shock Loadings Efficiently' was the heaviest weighted (35%) of any category or characteristic. This was a paramount concern because this was the main characteristic defined by our problem statement. 'Economic' Considerations category was the second heaviest weighted (30%). Cost-efficiency should always be one of the top priorities for good design. 'Operations and Maintenance' and 'Start-up Costs' were both assigned 20% weights. These weights are tied to the fact that we needed to be able to create a design that was feasible to test with the time and resources granted to us. More nominal categories such as aesthetics and implementation were also included in the design matrix with a small weight attached.

The scoring rubric shown in Figure 23 Design Matrix Scoring Rubric, found in the Appendix, was used to rate each alternative according score from 1-5: 1 being the poorest score, and 5 being the ideal score. These scores were normalized to reflect their associated weight, and summed, resulting in a total score out of 100 for each technology.

Alternatives Overview

The Fixed Film Reactor (FFR) scored higher than all other technologies assessed with the matrix. The main principle for a FFR is to provide a film that bacteria can colonize. The FFR has the following advantages: low maintenance requirements, continuous flow capabilities, and high removal efficiencies.

The Sequencing Batch Reactor (SBR) scored second highest compared with all other technologies assessed with the matrix. SBRs treat wastewater such as sewage or output from anaerobic digesters in discrete batches. The SBR has the following advantages: mature technology, simplicity of implementation. The SBR scored highest in the category of Design Implementation. The SBR scored lowest in the economics section, because it requires a large operating and startup cost.

Our two lowest scoring alternatives were the Anaerobic Fixed Bed Reactor (AFBR) and Anaerobic Sludge Blanket (UASB). Both alternatives scored the lowest in the Treatment category. Anaerobic systems do not handle shock loadings well: whether they be hydraulic or concentration based. Anaerobic systems also require a knowledgeable operator for system management which resulted in a low overall O&M score. At this point, strictly anaerobic systems were dismissed as potential alternatives for our design purposes.

Final Selection

The FFR and SBR scored almost evenly in our design matrix, with the FFR scoring slightly higher. Both systems provided efficient BOD and TSS removal, as well as the ability to handle highly variable flows and BOD concentrations.

However, while the SBR scored the highest in the design implementation category, it scored lowest in the economic considerations category. The FFR scored consistently well in more categories. The downsides to the SBR include long startup times and high startup costs. Furthermore, batch systems can cause issues with flow management when a constant flow is required by the AdvanTex system. SBRs would require a larger reactor volume and two reactors so one could store incoming flow while a batch is being treated in the other.

Ultimately, we decided to move forward with a Moving Bed Bioreactor (MBBR). The MBBR is a variation of the FFR in which the fixed film is attached to mobile “carriers”. The advantages of the MBBR over the SBR include ease of microbe regeneration, plug flow characteristics, reduced sludge wasting, low maintenance, more compact design, lower capital cost, and infrequent system adjustments

Methods

Prototype Constraints

Design of the prototype was constrained by the materials available to us, a budget of $500, limited lab space and shop time to build the prototype using shop tools. Some materials were made available to us by the Biological and Ecological Engineering department and others were purchased with our budgeted money.

Prototype Design

To simulate the geometry of the full-size reactor a vertical column reactor was chosen for the prototype. The acrylic column reactor from a previous experiment was made available by the department for the study. Also provided were influent and effluent ‘muck tubs’ that held our pre-mixed influent batches and received our treated effluent. To provide our reactor with influent we used a peristaltic pump. Carriers were added during operation until an ideal mixing regime was achieved. The liquid volume of the reactor was five gallons and the media fill ratio was adjusted to 60%. That ratio equated to three gallons of media and a total moving-bed bioreactor volume of 5.6 gallons. A detailed schematic of the prototype is shown in Figure 4 show a static water of 12” controlled by decanter height.

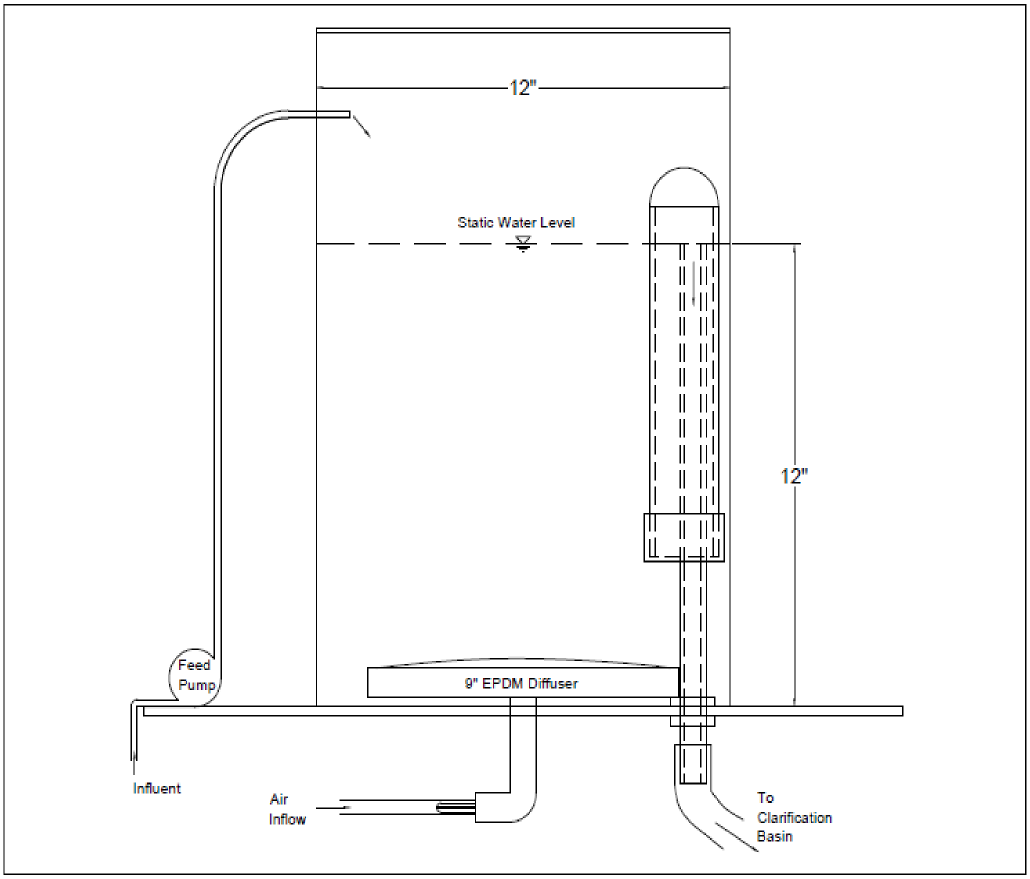

FIGURE 4 DETAILED MBBR SCHEMATIC

A sedimentation basin was designed to remove TSS from the effluent. A 2-gallon container with dimensions of 12"x6"x10" was used to model this basin with an actual fill volume of 2.06 gallons. Treated wastewater entered the sedimentation tank at one end, and excited at a set height at the other end. The hydraulic retention time inside the sedimentation basin ranged from 0.26-1.72 days (this variation in HRT was caused by peristaltic pump failures). The clarified and treated water then discharged to a muck tub.

The lab-scale prototype was run as a continuous flow reactor. A timer controlled, 12 V, peristaltic pump was used to control the flow of influent. Air was delivered to the system using a single 9" EPDM micro-bubble diffuser connected to a diaphragm pump with a flow rate of 793 gph.

Prototype Assembly

The prototype was constructed in the Oregon State University Open Source Environmental Sensing Laboratory and workshop under the supervision of Dr. Ganti Murthy, Dr. Chett Udell, and the staff of the OPENS lab. Assembly of the prototype began Wednesday, January 19th, 2017. The physical assembly period culminated on Wednesday, February 8th, when the completed physical prototype was transferred to the greenhouse for inoculation.

Prototype Components

A detailed list of the parts for the prototype is included in Table 1 below.

TABLE 1 PROTOTYPE COMPONENTS LIST

Reactor Vessel
Custom Plexiglas Lid
JB-Weld
Liquid Epoxy
Silicone
9" EPDM Diffuser
Two-way Splitter
Eco-Plus (793 gal/hr, 120 V) Aeration Pump and Tubing
Peristaltic Hydraulic Pump and Tubing
16 Gallon Muck Buckets x2
3 Gallons of Carriers
Sedimentation Basin and Tubing
Arduino Uno Microprocessor
Adafruit Assembled Data Logging shield for Arduino
MYPIN Waterproof Stainless Steel PT100 RTD Thermocouple Thermistor Sensor Probe
Adafruit PT100 RTD Temperature Sensor Amplifier - MAX31865
Gravity: Analog pH Sensor / Meter Pro Kit For Arduino
Perf Board, Various Wires, Solder

Data-logging Arduino

An Arduino Uno was used in combination with a data-logging shield (equipped with an RTC and an 8 GB SD card), an analog (industrial grade) pH sensor, and a digital (SPI connected) RTD temperature sensor. The Arduino was programmed to record and log both pH and temperature readings from the sensors.

Prototype Testing

Synthetic Waste Recipes

Synthetic waste recipes were initially 500ppm COD and incrementally increased to 5000 ppm COD. This was done so the microbial population could acclimate to concentrated loads. When unreasonable COD results came back from the lab, the recipe was adjusted again. It was determined that ethanol contained in the original recipe was evaporating during COD analysis, leading to lower than expected values in the influent category. Recipes used over time are summarized in Table 2 below.

TABLE 2 SYNTHETIC WASTE RECIPE SCHEDULE [30 L]

Testing Day	Target COD [mg/L]	Ethanol 95% [ml]	Sucrose [g]	Citric Acid [g]	Urea [g]	DAP [g]
1-2	500	7.5	2.4	None	0.65	1.4
3-4	1000	15	4.8	0.03	1.3	2.8
5-6	2000	30	9.6	0.03	2.6	5.6
7- 15	5000	75.4	23.7	0.03	6.5	14
15 -End of Testing	5000	0	139.2	2.0	6.5	14

Nutrient additions were calculated to maintain a ratio (C:N:P) of 100:5:1. Sucrose, ethanol and citric acid additions calculated to maintain the target BOD according to the standard 30 L batch recipe provided by our instructors. Sucrose was used until day 15 when the new recipe used dextrose was implemented. Table 3 below is a breakdown of the COD composition of the synthetic waste recipe.

TABLE 3 COD COMPOSITION

Component	Composition
Sucrose	COD 1.123g O2/g Sucrose
Ethanol	COD 2.182 g O2/g Ethanol
Citric Acid	COD 0.75 g O2/g Citric acid
pH	6.5
Temperature	Ambient

Chemical Oxygen Demand (COD)

Samples for COD analysis were collected from *the newly mixed influent batch* daily at 8 AM and from the reactor at 8 AM and 8 PM daily. Samples were taken using a 1 mL calibrated pipette. Samples were then pipetted into glass vials containing sulfuric acid and dichromate. They were then labeled to note the team, week, day, cycle, and replicate number. Sample vials were collected and taken to the department lab for digestion and analysis. The calibration curve used for spectrophotometer analysis and COD content is included in the appendix.

Effluent samples were twice filtered using a repeated process of straining effluent through filter pads beginning March 3, 2017. Prior to this date samples were unfiltered. The filtration paper used had a clean flow rate of 5mL/min and particle retention of 1-3µm. Filtration became necessary as biomass accumulation in the reactor began to influence COD test results. Because the biomass itself is made of organic chemicals that oxidize the COD results appeared to show less removal than was occurring.

Dissolved Oxygen (DO)

Dissolved Oxygen was sampled with a Milwaukee Smart DO Meter. DO was measured in both the reactor and the influent twice a day, morning and evening. The DO meter was calibrated using a 0 DO solution of ferrous sulfate. DO was sampled twice daily in the MBBR reactor (assuming completely mixed conditions) and in the newly mixed batch of synthetic influent.

Microbial Respiration Rate (K_d)

Microbial respiration rates were measured with a series of handheld probe measurements. Reactor dissolved oxygen concentration was measured with a handheld probe to confirm that the concentration was stable. The air-supply was turned off and DO (C) concentrations were measured at timed intervals until the concentration reached a stable minimum. These tests were performed on March 2nd, 7th, 8th, and 10th. Results were plotted with a first-order decay assumption represented by the following equations:

$$\frac{dC}{dt} = -K_d * C \qquad \text{[Equation 1]}$$

$$K_d = -\frac{Ln\left(\frac{C}{C_O}\right)}{t} \qquad \text{[Equation 2]}$$

The slopes of the fitted lines were averaged and this average was used as the COD decay rate (K_d) in the final model. The critical assumption of these tests we're that rate of oxygen consumption was equal to the rate of COD consumption, dissolved O_2 concentration is the only rate limiting factor, respiration is represented as a first-order reaction, the system is operating at steady-state, and the reactor is a completely mixed system.

Oxygen Transfer Rate Coefficients (K_{La})

The oxygenation rate coefficients (K_{la}) were determined experimentally for the reactor with clean water (no microbes or carriers) and wastewater (microbes + carriers).

Wastewater

To determine the K_{La} of the reactor with wastewater, the air supply was turned off for 10 minutes, DO was measured to confirm the DO concentration in the reactor had reached a stable minimum. At this point the air-supply was turned back on and the DO was measured at timed intervals until the concentration reached a stable maximum. The K_{La} was calculated with the following equations:

$$K_{La} * (C^* - C^{max}) = K_d * C^{max} \quad [\textit{Equation 3}]$$

$$K_{La} = \frac{K_d * C^{max}}{(C^* - C^{max})} \quad [\textit{Equation 4}]$$

Where K_d is the experimental value collected from the oxygen consumption tests, C* is the saturation of DO [mg/L] at temperature [^oC], C^{max} is the maximum DO concentration [mg/L] reached in the reactor.

Clean Water

To determine the clean water K_{La} the reactor was emptied of wastewater and carriers, cleaned, and filled to the fill line with tap water. The DO in the reactor was brought to zero by adding a measured amount of Na_2SO_3 based on the measure DO concentration. When the DO had been brought to zero the air-supply was turned on and the DO was measured at 15 second intervals for 7 minutes. This test was repeated 3 times. The K_{La} was determined based on the average of the three sampling periods.

$$\frac{d(C^* - C)}{dt} = -K_{La} * (C^* - C) \quad [\textit{Equation 5}]$$

$$K_{La} = -Ln\left(\frac{(C^* - C)}{(C^* - C_o)}\right) / \, t \quad [\textit{Equation 6}]$$

Temperature Adjustments

For further calculations, all measured K_d and K_{La} values were converted to values at a temperature 20 ^oC (K_{d_20} & K_{La_20}) with the following equations:

$$K_{d\_20} = \frac{K_d}{\Theta_d^{(T-20)}} \quad , \quad \Theta_d = 1.048 \quad [\textit{Equation 7}]$$

$$K_{La\_20} = \frac{K_{La}}{\Theta_a^{(T-20)}} \quad , \quad \Theta_a = 1.024 \quad [\textit{Equation 8}]$$

These conversions can also be used to predict the treatment capability at various temperatures to determine the influence of seasonal and climatic changes on efficiency.

Modeling

Using the measured clean water K_{La}, and the measured K_d, a modeled reaeration curve was plotted against the sampled DO values from the March 2nd test. This model was generated using the following equation. The full transient solution is derived in Appendix A2.

$$\frac{dC}{dt} = K_{La} * (C^* - C) - K_d * C \qquad [\textit{Equation 9}]$$

Ratio of Wastewater to Clean Water K_{La}

The mean value of the experimental K_{La} value from the clean water tests and the mean value of the experimental waste water K_{La} were compared to determine the impact of the waste water and carriers on oxygen transfer. This value can be referenced for aeration requirements in the full-scale design. The ratio is determined in the following equation:

$$\alpha = \frac{Wastewater\ K_{La}}{Clean\ Water\ K_{La}} \qquad [\textit{Equation 10}]$$

Temperature

Temperature was sampled and recorded twice daily. The ambient temperature was taken using a digital wall thermometer. The influent and reactor temperature were sampled using a digit rod thermometer probe as well as being continuous recorded every minute using a General-Purpose thermometer probe that was read and collected by an Arduino Uno.

Total Suspended Solids (TSS)

Total suspended solids were analyzed with a single sampling event on March 10, 2017. Samples were collected from both the main reactor and the sedimentation basin to determine the settling rate. Prior to sampling filter papers with a fine porosity capable of filtering 1-3µm particulate were placed in a desiccator to remove any moisture for 24 hours and weighed. 45 ml samples were pulled through the dried filter papers and placed in an oven at 110 C for 2 hours.

pH

pH was sampled twice daily in the influent and reactor using an Oakton EcoTest pH 2 waterproof probe with a range from 0 to 14. In addition to the probe, a continuous testing DF robot SEN169 pH sensor in the reactor was linked to an Arduino Uno.

Full Scale Design

Tank Sizing

The tank is sized based on the following equation to allow proper hydraulic retention time. This equation assumes that the full-scale reactor will have the same COD consumption rate as our lab test and that in our experiment chemical oxygen demand (COD) was equal to biochemical oxygen demand (BOD). This is a reasonable assumption as all the substrates in our feed compound were readily biodegradable compounds. Additionally, it is assumed that the reactor is completely mixed with all elements having equal retention times and that the system is operating at steady-state.

$$BOD = \frac{BOD_{in}}{1+K_d*\tau} \qquad [\textit{Equation 8}]$$

$$\tau = \frac{\left(\left(\frac{BOD_{in}}{BOD}\right)-1\right)}{K_d} \qquad [\textit{Equation 9}]$$

$$\tau = V/Q \qquad [\textit{Equation 10}]$$

Blower Sizing

Oxygen Requirements

Aeration requirements are based on the daily maximum loading of BOD that the system will be expected to remove.

$$AOR_{max}\left[\frac{lb_{O2}}{min}\right] = BOD\left[\frac{mg}{L}\right] * Q[gpd] * 3..79\left[\frac{L}{gal}\right] * \frac{1}{10^6}\left[\frac{kg}{mg}\right] * 2.2\left[\frac{lb}{kg}\right] * \left[\frac{1}{1440}\frac{day}{min}\right] \quad [\textit{Equation 11}]$$

$$AOR = AOR_{max} * C_{fine}$$

AOR_{max} = Actual Oxygen Required with maximum BOD loading

C_{fine} = 1.10, adjustment factor for fine bubble diffuser system [13] (Red Valve, 2002)

The calculations for required blower size were performed with the following equation.

$$SCFM = \frac{AOR_{max}}{\alpha} * [SOTE * H] * 0.0173\left[\frac{lbO2}{SCFM}\right] * FS \qquad [\textit{Equation 12}]$$

SOTE = 2.05% /ft = Assumed standard oxygen transfer efficiency in clean water for fine bubble diffusers. [13] (Red Valve, 2002)

H = depth of water column [ft]

FS = Factor of safety = 1.2 [13]

Max Duty

In addition to the SCFM output required by the blower, it is important that the blower have enough available head to pump against the pressure of the water column as well as the frictional head losses created by the diffusers. This Max duty is reported on blower performance in units of inches H_2O.

Economic Analysis

Assumed Motor Efficiency: Ɛ = 70%

Brake Horse Power:

The following equation quantifies the brake horsepower measured just outside the crankshaft, this is before any losses due to power allocations to other parts of the blower.

$$BHP = HP * \mathcal{E} \qquad [\text{Equation 13}]$$

Estimated kW of motor:

The power conversion from brake horsepower to kW was used to estimate the rate of energy transfer from the blower.

$$kW = BHP * 0.746 \frac{kW}{BHP} \quad \text{[Equation 14]}$$

Power:

Conversion from kWh to $ used to estimate the annual electrical cost associated with running the blower.

$$kWh = kW * hr * days \quad \text{[Equation 15]}$$

$$Total\ Cost = kWh * \frac{\$}{kWh} \quad \text{[Equation 16]}$$

Results

Prototype Testing

COD Removal

The mean values for the triplicate samples were used in comparing influent and effluent COD values, in addition to computing the daily COD removal percentages in Figure 5 below, the relationship between percent COD removal and the hydraulic retention time within the reactor is shown.

TABLE 4 DAILY REMOVAL EFFICIENCIES

Time [Day]	COD Removal [%]
1	35.3
2	49.1
3	15.1
4	No Data
5	32.5
6	23.0
7	38.3
8 (Began Filtering)	**52.8**
9	**29.4**
10	**82.3**
11	**47.3**
12	**No Data**
13	**58.6**
14	**58.6**
15	**76.8**
16	**60.7**

Figure 5 shows a plot of the changes in COD removal and hydraulic retention times with time. As our project progressed we had an increasing number of pump failures leading to increasing hydraulic retention times. As HRT increased, treatment efficiency would be expected to increase as well. This relationship was not observed as expected. It appears that removal efficiencies did improve with time as our reactor developed, but further tests would need to be performed with greater control of the hydraulic retention time to confirm this.

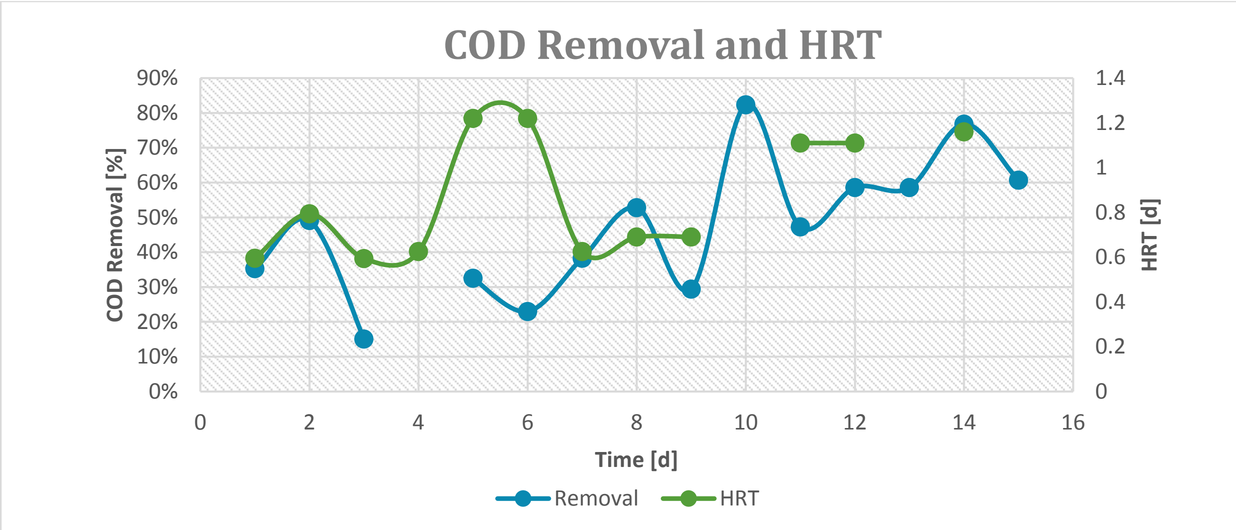

FIGURE 5 GRAPH OF COD REMOVAL EFFICIENCIES

Figure 6 shows a plot of COD removal (%) compared to the hydraulic retention time in the reactor. From this plot it appears that there was no significant correlation between HRT and COD removal during our experiment. Further tests would need to be performed to determine the cause of this.

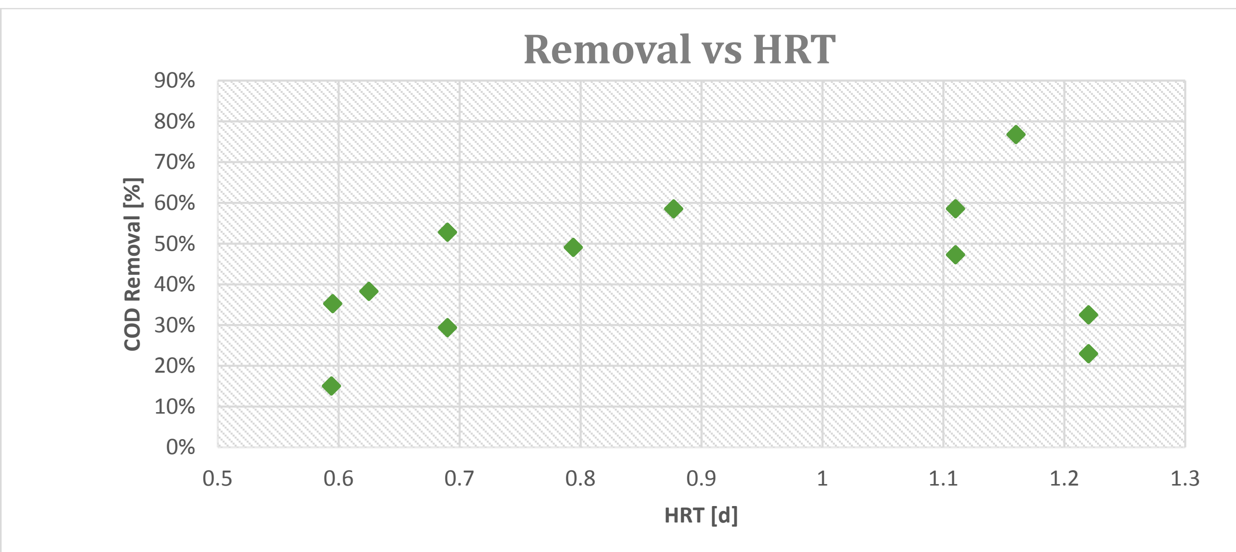

FIGURE 6 GRAPH OF REMOVAL EFFICIENCIES AGAINST HYDRAULIC RETENTION TIME

Dissolved Oxygen (DO)

Dissolved oxygen levels are shown in Figure 7. The mean DO level in the effluent was 6.0 mg/L, and 7.9 mg/L in the influent. The decrease in the dissolved oxygen is due to the microbial degradation of COD in the reactor.

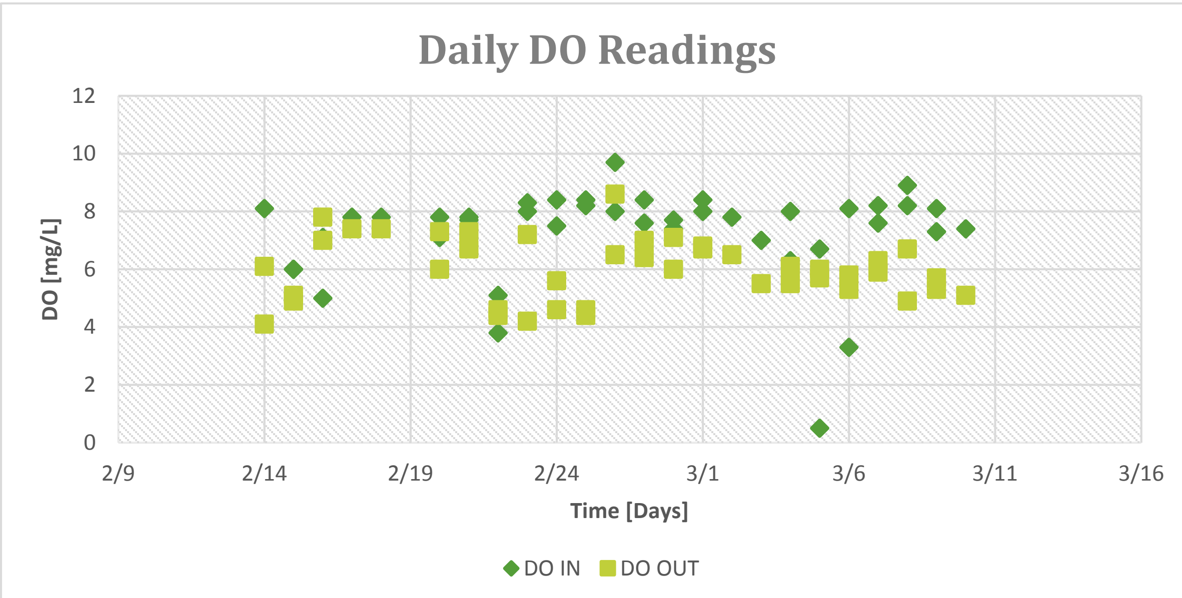

FIGURE 7 INFLUENT AND REACTOR DAILY DISSOLVED OXYGEN READINGS

Microbial Respiration Rate

Measurements of microbial respiration rates were collected on three separate days. The plots of the measured DO concentrations vs. time are shown in Figure 8. This shows similar rates of decay from each of the three sample periods.

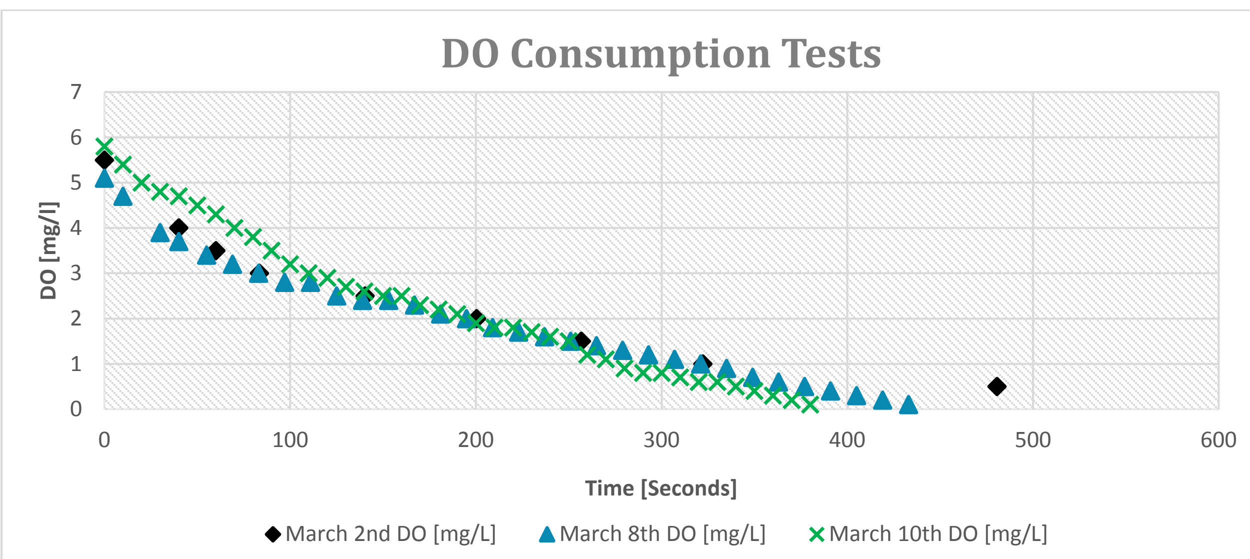

FIGURE 8 DISSOLVED OXYGEN TESTS

To test the assumption that the DO consumption rates act as a first-order process, the data from the three sample periods were plotted as the natural log of the fraction of instantaneous DO and the initial DO versus time. The results of this are shown in Figure 9. The data collected on March 2nd showed a near perfect first-order relationship; however, the data from the sampling periods on March 8th and 10th appear to show a more linear relationship.

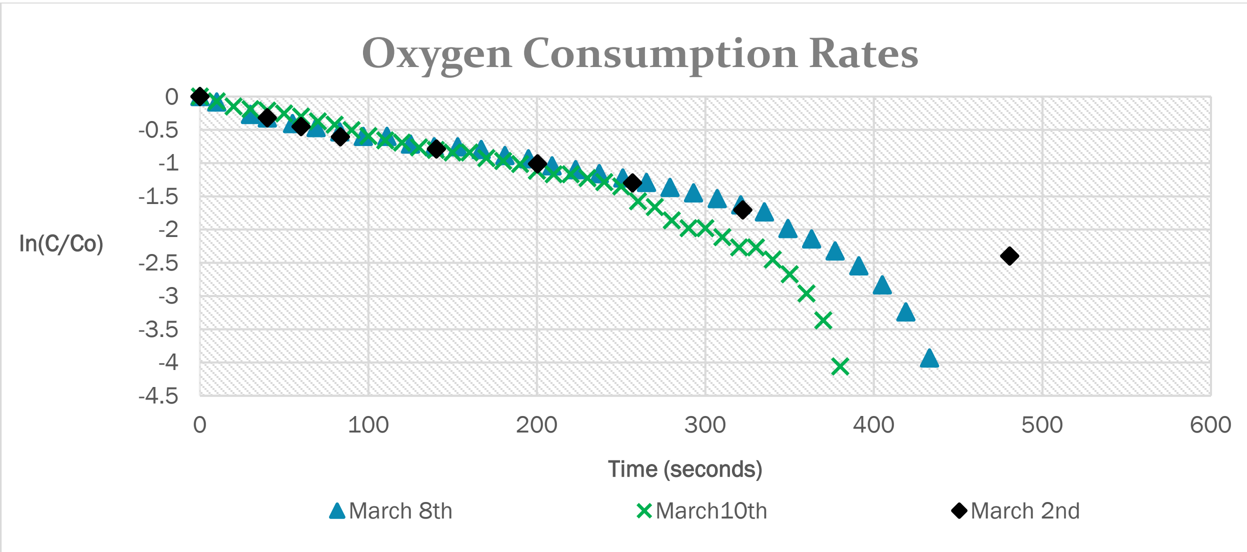

FIGURE 9 FIRST ORDER KINETICS OF OXYGEN CONSUMPTION

The first-order decay constants obtained from are summarized below in Table 5. The average first-order decay rate obtained from the sampling periods, using equation 1, was 0.0065 s^{-1}.

TABLE 5 FIRST ORDER KINETICS OF OXYGEN CONSUMPTION AND CORRESPONDING R^2 VALUE

Test Date	K_d [s^{-1}]	R^2	Reactor Temperature [C]
March 2nd	0.0048	0.9927	18
March 8th	0.0067	0.8799	18
March 10th	0.0081	0.9018	18
Average	0.0065	-	18

Oxygen Transfer Rate Coefficients (Kla)

The linearized results of the three clean water aeration tests are shown in Figure 10. The rates derived from this plot are summarized in Table 6.

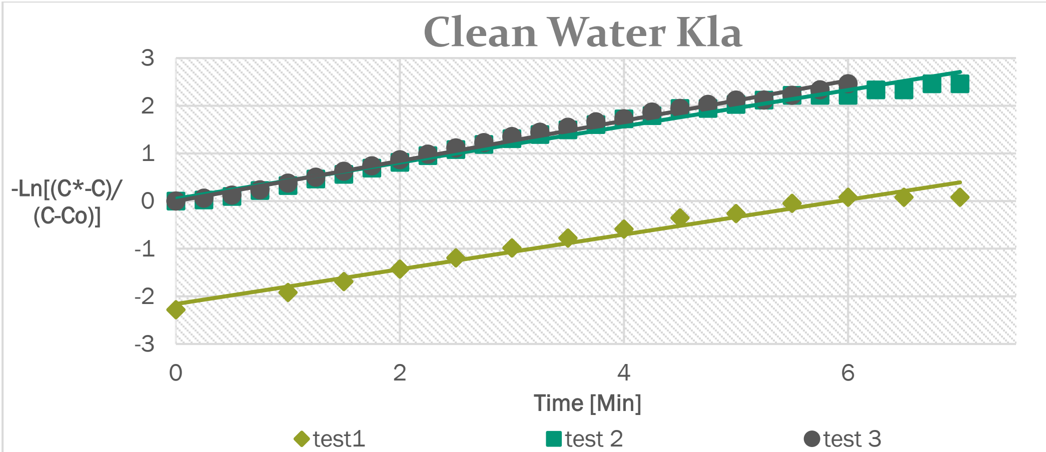

FIGURE 10 GRAPH OF CLEAN WATER KLA

TABLE 6 CLEAN WATER KLA TEST

Test	K_{La} [s^{-1}]	R^2	Reactor Temperature [°C]
1	0.0070	0.9935	22
2	0.0063	0.9794	22
3	0.0061	0.9718	22
Average	0.0065	-	22

The linearized sample data from the wastewater aeration test on March 2nd is shown in Figure 11. This data appears to show less of a first-order relationship than the clean water tests, indicating that other variables may be impacting the aeration rates, most likely this variable is the microbial respiration rate which is not present in the clean water tests.

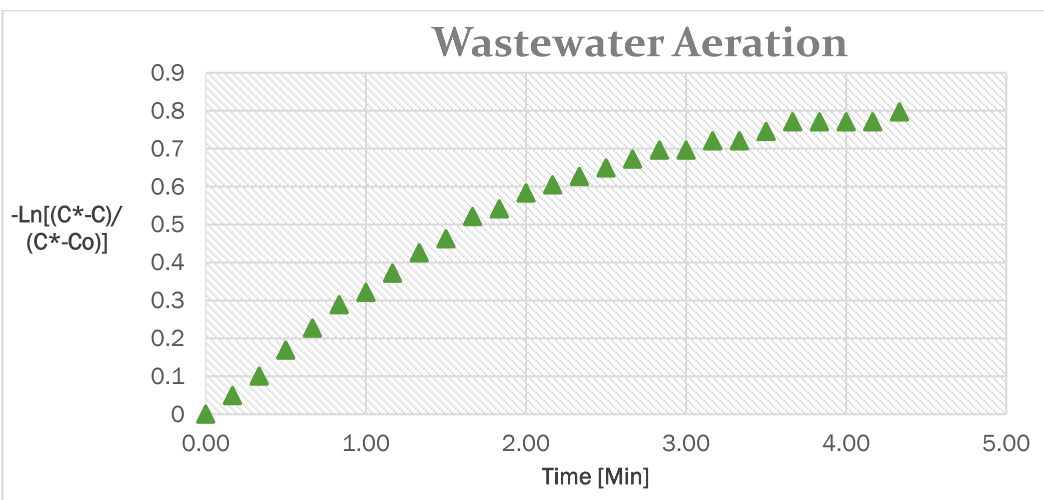

FIGURE 11 WASTEWATER AERATION

From the oxygen consumption rates summarized in Table 5 the K_{La} rates for the wastewater could be determined with equation 4. The calculated wastewater K_{La} values are summarized below in Table 7.

TABLE 7 WASTEWATER KLA VALUES

Test	K_{La} [s^{-1}]	DO [mg/L]	Reactor Temperature [°C]
March 2nd	0.0071	5.6	18
March 8th	0.0080	5.1	18
March 10th	0.0130	5.8	18
Average	0.0094	5.5	18

Using the average wastewater K_{La} from Table 7, and the average K_d from Table 5, a model was developed using equation 9. To test this model the modeled oxygen concentrations were plotted along with the sampled values from the March 2nd sampling period. Additionally, a model was plotted using the mean clean water K_{La} value to show the comparison with the wastewater K_{La} values. The results of these models are shown below in Figure 12.

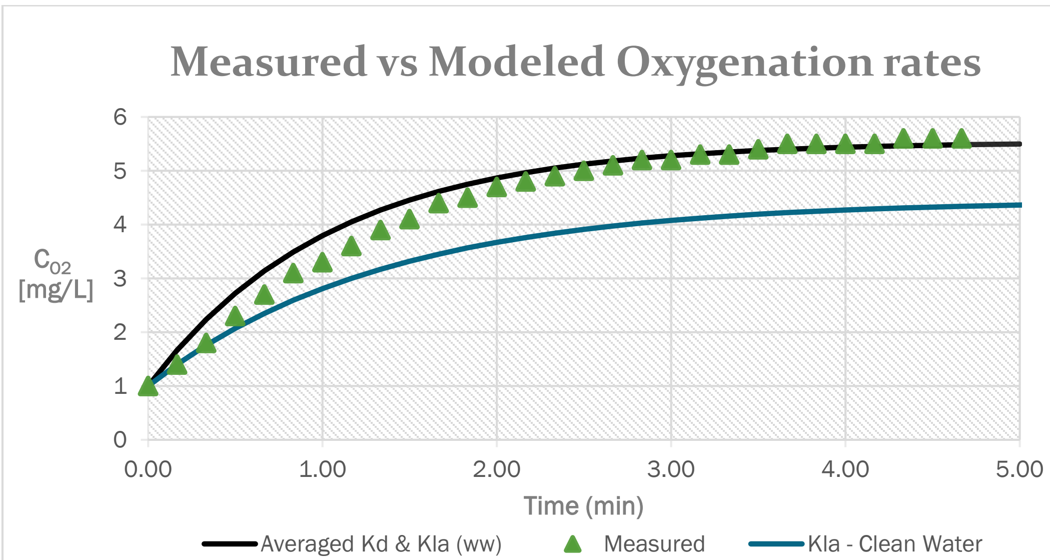

FIGURE 12 MEASURED AND MODELED OXYGENATION RATES

For the model using averaged experimental K_d and K_{la} values the RMSE = 0.249, for the clean water K_{la} RMSE = 0.974. This shows that there is a significant change in K_{La} between the wastewater and clean water within this reactor.

The determined coefficient for conversion of clean water to wastewater K_{La} using equation 10.

$$\alpha = 1.6$$

Total Suspended Solids (TSS)

The sample taken from the reactor showed a TSS of 1100 mg/L. The sample taken from the sedimentation basin did not have a measurable TSS. Filtered water from both samples remained cloudy after two passes through the filter.

pH

pH levels were variable and fluctuated between 6.4 and 7.9. The variation can be seen in Figure 13 and Figure 14 shows an oscillating pattern.

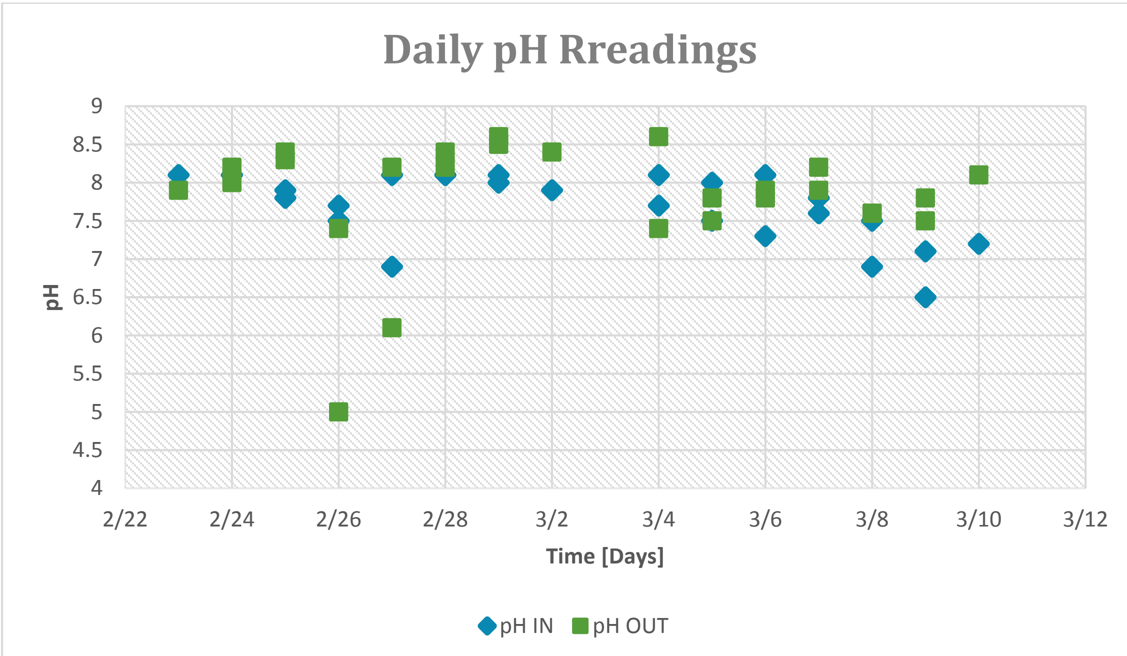

FIGURE 13 INFLUENT AND REACTOR PH DAILY READINGS

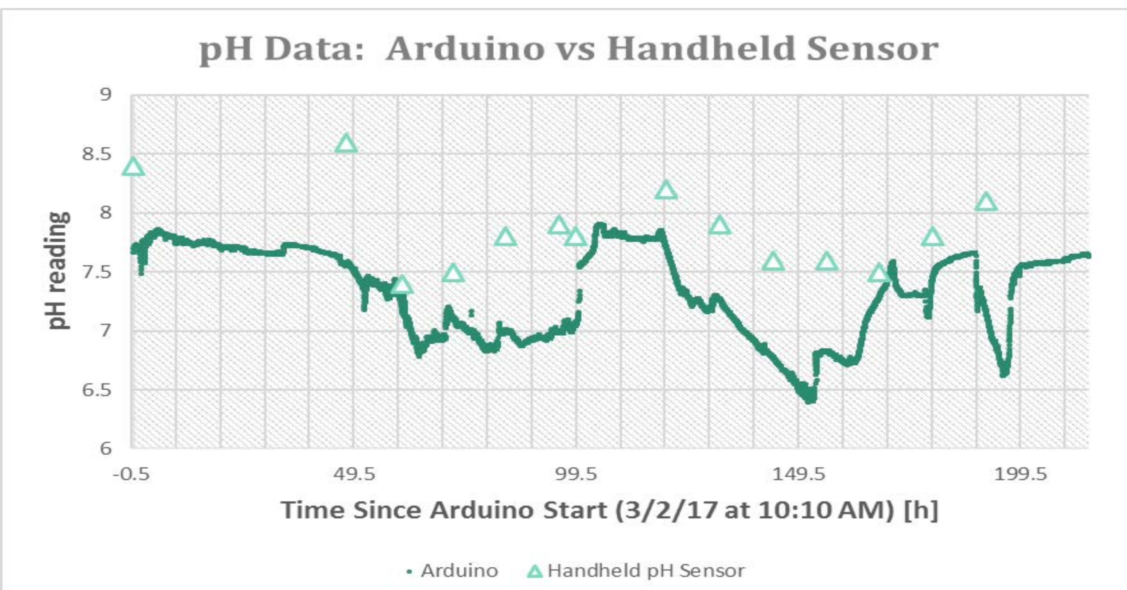

FIGURE 14 COMPARING ARDUINO PH TO GREENHOUSE HANDHELD SENSOR

Temperature

On average, the reactor temperature was 2.9 °C warmer than the influent, likely due to a combination of biological activity, and warm air from the pump. The influent temperature was comparable to the ambient temperature inside the greenhouse. The temperature inside the greenhouse is moderated by a temperature control system, but was also highly influenced by the weather and temperature outside. Figure 15 and Figure 16 show the general-purpose thermometer and the Arduino readings.

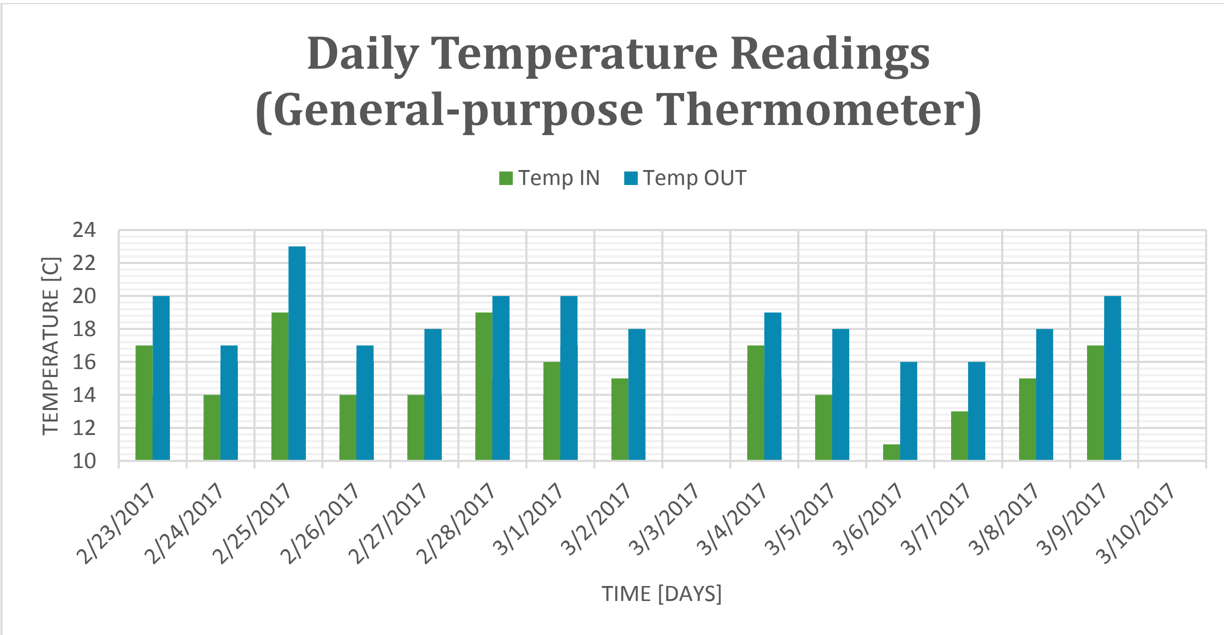

FIGURE 15 INFLUENT AND REACTOR DAILY TEMPERATURE READINGS

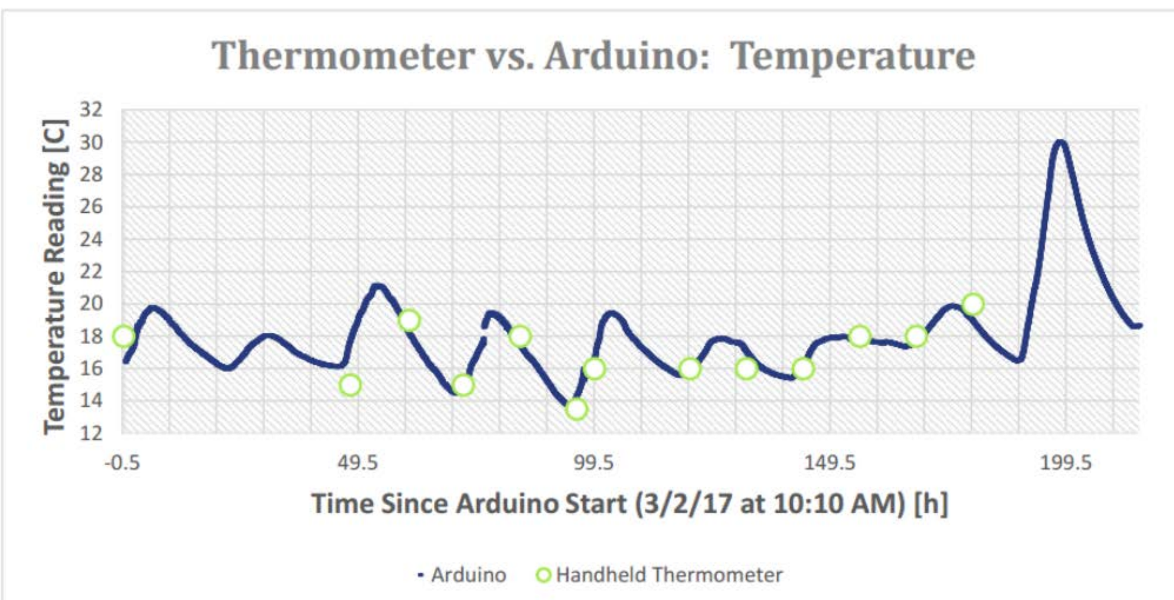

FIGURE 16 COMPARING ARDUINO TEMPERATURE TO GREENHOUSE GENERAL-PURPOSE THERMOMETER

Figure 17, below, shows a plot of recorded dissolved oxygen vs temperature inside the reactor.

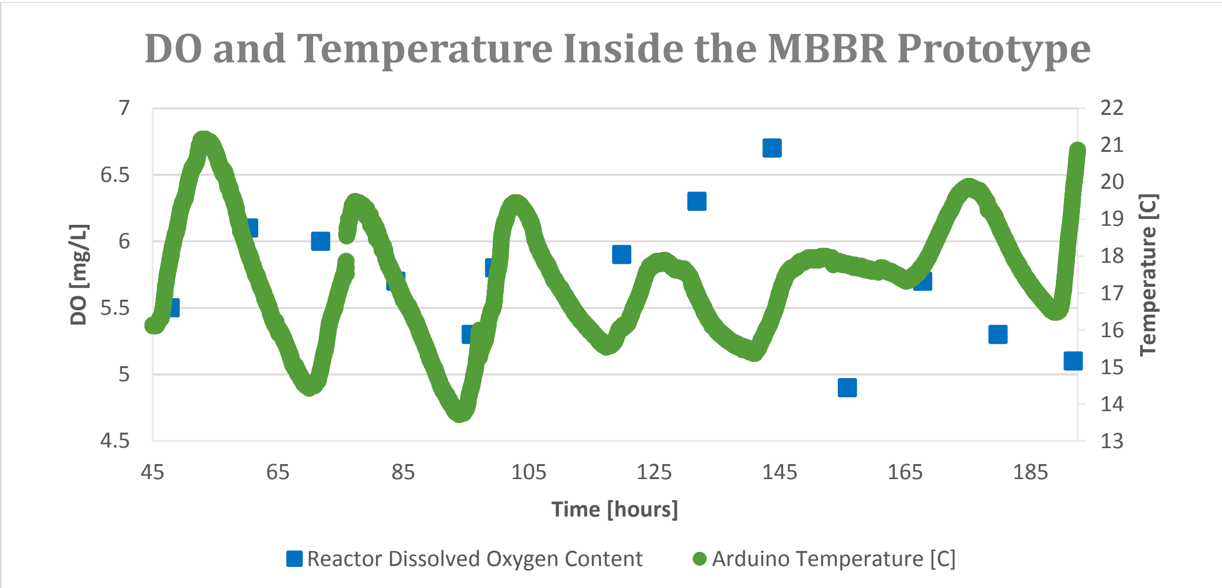

FIGURE 17 COMPARING DISSOLVED OXYGEN IN REACTOR TO TEMPERATURE

Discussion

Analysis

The peristaltic pump continually failed to deliver a consistent 8 gallons per day (gpd) which created the large variation in hydraulic flow.

COD removal rates were variable, but showed an increasing trend over the experiment period. This can be attributed to the bioreactor still developing. The increasing microbial biomass on carriers would eventually lead to greater removals of COD.

Figure 21 shows the carriers with microbial biomass growing within the interstitial areas.

Another reason for the increase in COD removal percentage was a switch in sampling protocol that began on day eight of the sampling period. This change in protocol was to add a double filtering process before effluent was added to reagents in the analysis tubes. After this change, there was a sharp increase in COD removal. This can be observed in Table 4. The increase in the removal percentage reported is due to the unfiltered effluent containing a significant mass of suspended solids in the form of microbes which showed up as COD during analysis.

The prototype MBBR showed consistent removal over the final four days with an average of 64% of the COD being removed. The average COD removal after the filtration protocol was implemented was over 58%. Based on this data alone this reactor did not have a significant enough removal rate to treat maximum expected concentrations of 7000 mg/L down to 500 mg/L or less. An upward trend in removal was seen over the testing period. This leads us to believe that the MBBR may have shown better removals if there had been more time for the reactor to acclimate and for the carriers to increase their attached biomass. Another option to increase removals is to use reactors in series. This would improve treatment efficiencies in the full-sized design. Time, budget and space limitations prevented these comparisons.

The model developed for the averaged K_{La} and K_d values showed a good fit with an RMSE of only 0.249. This showed a ratio of wastewater to clean water K_{La} (α) of 1.6. This value is higher than the expected range of 0.4 – 1.0, however studies have shown there to be an increase in K_{La} for MBBRs when compared to the clean water due to the influence of the suspended carriers on the residence time of the bubbles [12]. Further tests would have to be performed on a full-scale reactor to confirm this value. Additionally this value could be heavily impacted by compounds present in actual winery wastewater that are not present in the synthesized waste fed to our reactor.

The measured microbial respiration rates were far higher than those predicted by our sampled COD values. Our sampled values showed that even with a residence time of 1.2 days our removal percentages still only reached a maximum of %84. According to our averaged handheld measurements of respiration rates shown in Table 5 our removal efficiency should have approached more than 99%. This would have to be experimented with further, but we are hypothesizing that this difference was present due to the high amount of COD present within the suspended biomass. This was apparent in the average COD reduction of 50% between samples which were filtered compared to those that were not filtered. Measured TSS was 1100 mg/L and after two passes through a 3 µm filter the water was still very cloudy indicating large amounts of fine suspended biomass. This suspended biomass may have still represented a significant source of COD. This could be further removed through chemical flocculation.

The DO levels inside the prototype reactor stayed well above the necessary 2.0 mg/L that microbes need to degrade organic compounds [1]. This shows that oxygen was not a limiting factor in the reactor system. For scale-up, the system should be designed to meet, but not overly exceed the 2.0 mg/L of dissolved oxygen inside the reactor. This would meet the removal efficiency and reduce the amount of needed input energy, reducing annual costs.

Removal efficiencies and hydraulic retention times showed little correlation. We would expect to see a strong correlation with longer retention times equating to higher removal because HRT has a direct effect on bioreactor treatment efficiencies.

The Arduino Uno and pH meter recorded pH over time showing an oscillating pattern. This displays the self-buffering capacity of the MBBR system. If there are large changes in the influent's pH, then the system should self-adjust and should naturally fluctuate between 6.5 and 8.0. The prototype showed a range in pH of 6.4 to 7.9. However, if large concentrations of acids or bases are introduced to the influent through the winery's use of cleaning agents, a pH buffering system would be needed. This is an aspect of the design that would be based upon individual winery's needs and could manually be adjusted by winery staff based on the process and chemicals being used.

Temperature fluctuations in the influent and treated effluent were highly dependent on the daily weather and air temperatures. The prototype experiment was in a university greenhouse with a temperature maintained at between 10 – 20 ^{o}C. The influent temperatures were generally close to the ambient air temperature, but the reactor's temperature averaged 2.9 ^{o}C more than the influent. This can be attributed to the energy of the motion inside the reactor and the energy output of the microbes. No correlation was observed between the DO concentrations and the temperature inside the reactor. We would expect a strong correlation between the two variables, but there were problems in getting accurate and dependable DO data due to the storage and improper use of DO meter. Another possibility in not seeing the correlation between the reactor temp and DO in the reactor is that the variation in the temperature was only 5 ^{0}C with much of the variation being within 3 degrees. The variance in maximum saturation concentrations of dissolved oxygen at these temperatures is only 1.0 mg/L. This may account for the lack of a strong correlation.

Effluent data from the day 12 of sampling was excluded from the analysis due to a large volume of influent accidently being pumped through the reactor previously that day. This significantly reduced the hydraulic retention time and treatment efficiency skewing the evening data.

Error Considerations

Methods

To prevent variations in COD values, influent samples should be more thoroughly mixed. A better protocol for influent mixing could have been developed to ensure that the sugars were thoroughly dissolved and that the solution had reached equilibrium.

Initially the COD constituents were sucrose, citric acid powder and ethanol. When samples were heated during digestion and analysis, the ethanol was partitioning from liquid to gas phase. This caused the ethanol in the samples to not be reported as COD. A switch in influent batch mixing protocols was implemented to solve the problem. This could have been prevented with a more thorough analysis of COD constituents. Dextrose and citric acid could have been used from the beginning of the sampling period.

Equipment

The peristaltic pump that was used for the experiment had issues delivering a constant volume of influent and most of these problems may be attributed to the wrong tubing being provided and used. This could have been prevented with correct tubing material and diameter.

Final Recommendation

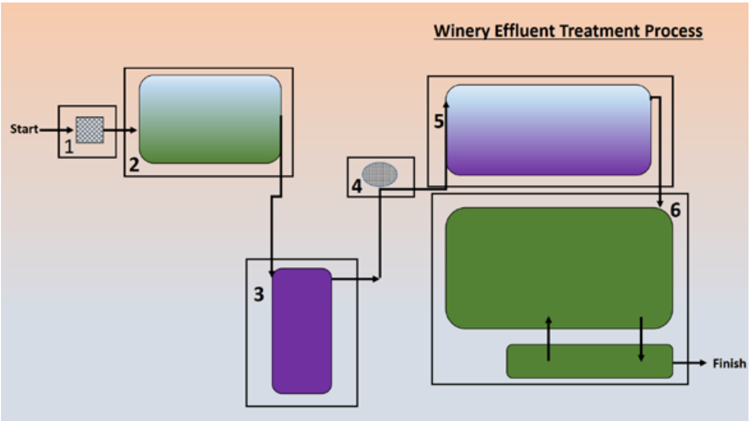

FIGURE 18 WINERY EFFLUENT TREATMENT PROCESS

The overall treatment train employed by OSI will involve the five main steps as shown above in Figure 18 and outlined below in Table 8.

TABLE 8 PRIMARY TREATMENT TRAIN STEPS

Step 1— Pre-Screen and Bio-augmentation	Pre-screening of the raw winery effluent occurs before entering the solids settling tank. pH is adjusted to be between 6-8.
Step 2—Solids Settling	The waste-stream enters a 3000-gallon solids-settling tank. A floating decanter, operating at 1000 gallons per day, will maintain a hydraulic retention time of approximately 1 day, leaving 1500 gallons as a surge buffer to protect against excessive flows.
Step 3—Primary MBBR Treatment	The waste-stream is pumped into the MBBR reactor series. The reactors are self-regulating by decanter height and can be taken off line when crush season effluent flows drop.
Step 4—MBBR Effluent Screening	Pre-screening of MBBR effluent to remove additional TSS created in the reactor series.
Step 5—Secondary Settling	The waste-stream is pumped from the reactor to a clarifier designed to remove suspended bio-solids and prevent biofouling of OSI's AdvanTex AX100 filter pod.

System Overview

Moving Bed Bio-Reactor

Eco-energetics contribution to this system is the design of the MBBR (step 3, above).

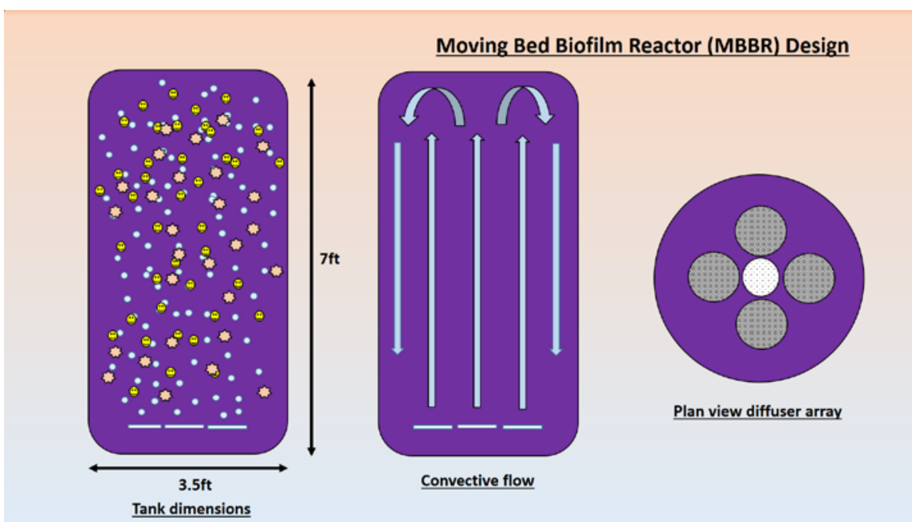

FIGURE 19 MOVING BED BIOFILM REACTOR DESIGN

The MBBR process shown above in Figure 20 will be incorporated into OSI's current treatment train as the primary treatment method. The purpose of this primary treatment system will be to reduce the BOD_5 of the incoming waste stream from a max of 7,000 mg/L to 500 mg/L or less prior to polishing in OSI's AdvanTex AX100 system. The MBBRs acting in series will result in greater BOD removals as the process is dependent upon concentration. As the concentration decreases the TSS in the reactors will also decrease, this has the benefit of a reducing the final effluent TSS from the third reactor. Figure 20 below depicts the flow through the MBBR tank.

To deal with the high loadings of TSS which are expected to be produced by the reactor a secondary settling tank will be required. Testing will have to be done to determine the need for chemical flocculation and the proper dosing and coagulant required for adequate treatment.

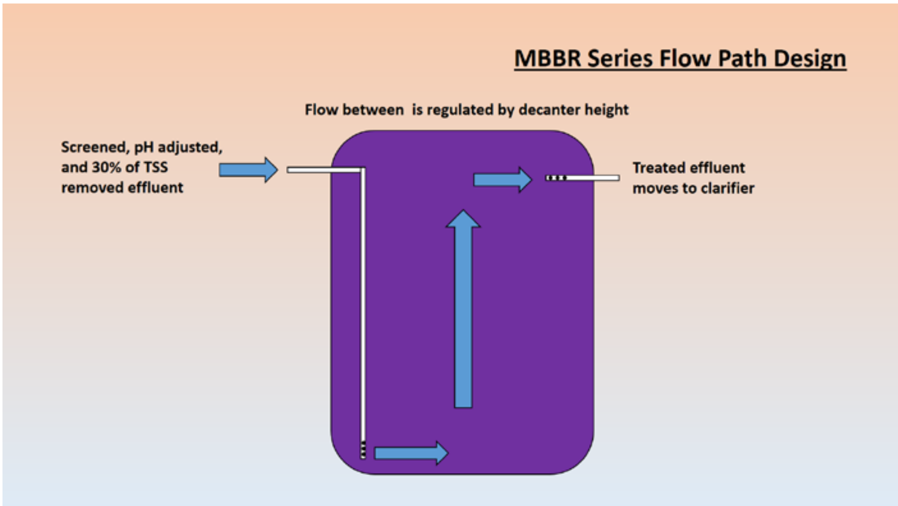

FIGURE 20 MBBR FLOW PATH DESIGN

Microbial Growth Carriers

The MBBR's will be filled to 60% (34.6 ft^3) with Dynamic Aqua Science AMB Bio-Media carriers. The AMB Bio-Media carriers are made of polyethylene and a have the following characteristics listed in Table 9 below.

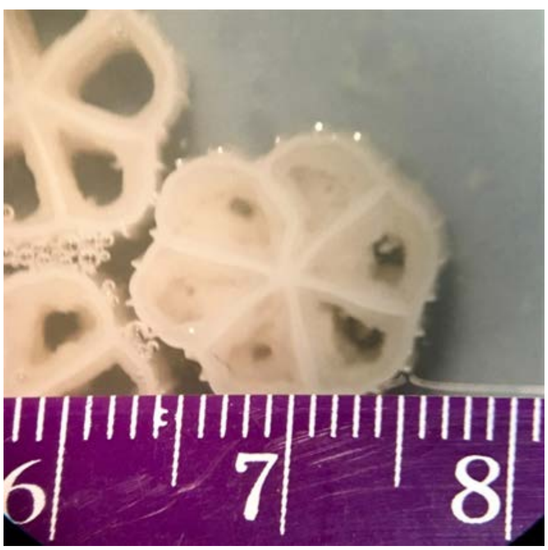

FIGURE 21 MICROBIAL CARRIERS

TABLE 9 MICROBIAL CARRIER SPECIFICATIONS

Dimensions	23mm x 17.5mm circle
Available bio-surface	400 m^2/m^3
Density at 23 ^oC	0.95-0.98 g/cm^3
Chemical Composition	HDPE
Service Life	20 years

The carriers provide microbes with a protected growth surface that is self-cleaning and has a large surface area to volume ratio. This maximizes effective surface area and maintains a high volume of biomass within the reactor, negating the need for sludge recycling.

Aeration Systems

Aeration is the most critical and costly component of any aerobic wastewater treatment system, typically accounting for 50-70 % of the energy used in the treatment process [5]. The aeration system must provide an adequate supply of oxygen to the aerobically respiring bacteria. Proper aeration maintains high BOD_5 removal efficiencies and prevents the system from becoming anaerobic. Without sufficient oxygen, the system could go anaerobic resulting in foul odors and longer retention times. Our calculations show that the proposed system will require a blower capable of providing 54 SCFM of air.

TABLE 10 AERATION COMPONENTS

Unit	Size	Bubble Size	Units Per Reactor	Cost Per Reactor
Pentair EPDM Disc Diffuser	Active 9” Total 10.9”	0.009” – 0.04”	4	$89.04
PermaCap Coarse Bubble Diffuser	Active 5” Total 5.3”	0.25” – 0.75”	1	$15

Diffuser Design

Aeration of the effluent is performed by four 9-inch EPDM micro bubble diffusers and one Environmental Dynamics PermaCap 5” coarse air diffuser (Table 10). The micro bubble diffusers will surround the coarse bubble diffuser in a diamond pattern (Figure 20). This configuration will produce a vertical “pumping” of effluent. Creating an updraft and maximizing turbulent mixing.

The four micro bubble diffusers will create small bubbles with a high surface area to volume ratio, with bubble sizes ranging from 0.009-0.04 inch. The oxygen transfer rate (OTR) will also increase (when compared to coarse bubble) due to the higher pressure within the micro bubbles. The coarse bubble diffuser located in the center will dominant mixing with the larger bubbles drawing fluid behind them as they rise, this creates a center roll mixing pattern.

Blower Design

The MBBR aeration system will be supplied by one Pentair AHPB70 regenerative blower (Table 11), featuring a torque controlling variable frequency drive controller (VFD). The VFD controller will allow the flow to be varied depending on the dissolved oxygen in the MBBR’s. This will maximize the efficiency of the MBBRs and cut aeration costs.

TABLE 11 BLOWER COMPONENTS

Unit	Output Horsepower	Power	Maximum head	Flow delivered	Unit Cost
Pentair AHPB70	3.4 HP	3.43kW	193 inches	48 – 63 CFM	$2084

Hydraulic Pumps

The primary sedimentation tank will feature a floating decanter capable of pumping 1000 gallons per day (Table 12). Another hydraulic pump in the ProSTEP™ pump package will deliver the pre-treated and clarified effluent to the AdvanTex filter pods.

TABLE 12 DECANTER COMPONENT

Unit	Flow Rate	Cost
Decanter Pump	4.2 [gpm]	$ 58.92

System Management

System management is designed to make operation as simple as possible and increase efficiency. Key system management components include automated control systems and biofilm upkeep.

Hydraulic Controls

Due to the highly variable and seasonal nature of the waste streams produced by wineries it is necessary to implement hydraulic controls. The MBBR system will be fed by the floating decanter in the primary sedimentation tank. After the mechanical pump in primary sedimentation, the reactors will self-adjust with decanters set at a permanent height that rely on the hydraulic head of the previous reactor to create a flow gradient. Each reactor will be fitted with a mechanical "max fill" float valve that will close the inlet valve on reactor one if any of the reactors in series experience an outlet failure. The wastewater will then accumulate in the primary sedimentation tank until the problem can be addressed. If the problem cannot be addressed in the allowable hydraulic retention time of the primary sedimentation tank a "max fill" float valve will close to the input flow. The shunted flow will then be rerouted to the pre-existing system.

These simplified controls do not require the use of fragile and expensive sensors which are prone calibration errors and electrical failures. This will reduce the operation and maintenance costs while maintaining adequate control of the systems physical operations.

Aeration Controls

The blower speed will be manually adjusted by a variable frequency drive (VFD) controller. This will allow the operator to adjust aeration based on measured DO concentrations in the reactor collected at sampling ports in each reactor with a hand-held galvanic DO probe. The pump speed and flow rate can be reduced to a minimum of 70% while maintaining adequate pressure. Additional reductions in airflow will be controlled through automated valves and a bleeder valve near the blower. This allows for complete control over airflow and a dramatic savings in energy costs.

Bioaugmentation

Seeding is recommended with 50% of the volume of all treatment tanks combined. The seed source should preferably be activated sludge (to favor aerobic microorganisms), which can usually be obtained from a municipal wastewater treatment plant. We recommend that to achieve maximum efficiency the reactor be given at least one month to fully develop to maximum BOD concentrations.

Biofilm Upkeep

The typical mean cell residence time (MCRT), the time that solids are retained on the carrier, for the average MBBR is 0.15 days [1] (note that this is the same as the Solids Retention Time or SRT). The benefit of the MBBR system is the carrier's ability to self-regulate the thickness of the biofilm layer to adjust for variations in organic loads. SRT is therefore inversely correlated to organic loading rates.

This ability to quickly self-regulate biofilm thickness also leads to increased levels of TSS in MBBR effluent. Cell slough from the growth media requires secondary clarification for effective BOD reduction. Sludge bulking can be prevented in aerobic processes by always keeping DO levels above 2 mg/L. The tank will require periodic cleaning which can be accomplished through vigorous mixing and removal through a valve in the bottom of each reactor.

Neutral pH Maintenance and Buffer Capacity

Neutral pH should be ensured by the pH control phase implemented by OSI. It is recommended that pH control mechanisms be outsourced to a third party.

Nutritional Requirements

Winery effluent is likely to be deficient in both nitrogen and phosphorus, both key macronutrients. The supplementation of these macronutrients is essential for effective winery wastewater treatment in an MBBR. Phosphorus deficiency can be moderately limiting for our system (3% reduction in BOD removal efficiency, on average), while Nitrogen can be especially limiting for our system (16% reduction in BOD removal efficiency, on average) [6].

Supplementation of these nutrients will be based on a COD:N:P ratio of 100:5:1 [10]. Based on this ratio, a supplementation of ~ 27 Nitrogen kg/year and ~5.2 Phosphorus kg/year will be needed [Appendix A6].

Temperature Maintenance

The thermal properties of fiberglass indicate that the temperature range will be acceptable year-round in temperate climates. However, it should be noted that the degree of sludge stabilization should be calculated at the lowest expected operating liquid temperature, while the maximum oxygen requirements should be calculated at the highest liquid operating temperature. This is a detail to our calculations that can be adjusted once we know more about the annual climate at the site in question.

Capital Cost Analysis

Listed below in Table 13 are the anticipated costs associated with the construction of the MBBR treatment series.

TABLE 13 CAPITAL COST ESTIMATE

Unit	Number	Cost Per Unit
Primary Sedimentation Decanter Pump	1	$58.92
EPDM Micro Bubble Diffuser	3	$22.26
Coarse Bubble Diffusers	1	$15.00
3.4HP 3-phase Blower	1	$2084.00
3-Phase VFD Controller	1	$240.00
High Water Alarms	2	$120.00
YSI Pro 20 DO Probe	1	$595.00
AMB Bio-Media Carriers	1	$600.00
3’x7’ FRP Reactor Tanks	3	$1700.00
Operations and Maintenance	Per/year	$665
Electricity Cost ($0.098/kWh)	Per/year	$1226
	Total Cost	**$4232**

Conclusion

Environmental Impact Considerations

The MBBR designed is a sustainable wastewater treatment technology. However, some wineries may use more energy to treat their wastewater onsite than they would to dump it directly into the municipal wastewater stream. If the treatment design can reduce the COD and TSS concentrations low enough, the winery can use the wastewater for irrigation purposes. Sludge can also be processed onsite to produce a high-quality compost.

Social Impact Considerations

The MBBR designed would have appealing aesthetics. The MBBR treatment train could be installed in a variety of places including underground or inside its own enclosed building. The design is an aerobic system, so when properly managed unpleasant odors would be kept to a minimum. If the winery chose, they could use the wastewater treatment system as part of the winery tour. Explaining the importance of proper wastewater treatment and increasing awareness of water treatment.

Economic Considerations

The MBBR design would have to be evaluated on a case by case basis for each individual winery. The costs of the winery's wastewater transport and or disposal would have to be analyzed and compared to the cost of design, installation, capital, and operational and maintenance costs to decide on whether the MBBR would be a cost-effective strategy.

Other Considerations

Feasible: The MBBR treatment train is feasible for small to mid-sized wineries and would work well with Orenco Systems Inc. AdvanTex system.

Practical: The MBBR System would require site operator experience and training. The installation of the MBBR system would be technically simple, but would require engineering design based on the specific winery's needs. The system is robust in the face of variable hydraulic and organic loading rates, has segmented design for variable flows, is self-regulating with respect to sludge production, and has a small footprint.

Economical: The cost to benefit of the MBBR system would have to be evaluated on a case by case basis.

Safety: The MBBR is an enclosed treatment system and with proper training in addition to process controls, the system is reliably safe.

Legal: Land regulations differ from region to region, but the MBBR system should be in compliance with all regulations. This would need to be checked in each case.

Moral: Using the MBBR system to reduce the BOD of the wastewater results in an effluent that will not contaminate or degrade the environment or affect human health in a negative manner.

Acceptable: The prototype MBBR system showed good removal of COD concentrations.

Ecological: The MBBR is an ecologically appealing wastewater treatment system.

Acknowledgements

This design would not have been possible without the consultation and insight of the following: Tristian Bounds, Head R&D Engineer with Orenco Systems Inc., Professors Ganti Murthy and Roger Ely of Oregon State University's Biological and Ecological Engineering department. Professors Dr. Brian Wood, Dr. Tyler Radniecki, Dr. Lewis Semprini of Oregon State University's Chemical and Environmental Engineering department, Professor Dr. James Osborne of Oregon State University Food Science and Technology

department, Dr. Chet Udell, and the staff of the Oregon State University Open Source Environmental Sensing (OPENS) Laboratory.

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Appendix

A1. Alternatives Analysis

Technical Alternative →	Weight	SBR	MBR	MBBR	JLR	TMBR	FBAR	UASB	RBBR
Economic									
Operating costs	10	8	6	6	8	6	6	8	8
Start-up cost	20	12	8	16	16	12	8	4	8
Overall Economic score	30	20	14	22	24	18	14	12	16
Treatment									
BOD Removal Efficiency	12	12	12	9.6	9.6	9.6	4.8	4.8	12
Handles variable flow	10	10	10	10	10	10	2	2	6
Handles variable concentrations	10	8	10	10	10	10	4	2	8
Retention Time	3	2.4	2.4	2.4	2.4	0.6	1.8	1.8	0.6
Overall Treatment Score	35	32.4	34.4	32	32	30.2	12.6	10.6	26.6
Design Implementation									
Complexity	5	4	2	4	3	4	4	4	3
Tech Maturity	5	5	3	4	2	5	4	4	3
Overall Design Score	10	9	5	8	5	9	8	8	6
Operation and mainetance									
Skill required	6	3.6	2.4	6	4.8	4.8	4.8	2.4	2.4
Material availability	4	4	2.4	3.2	3.2	3.2	2.4	2.4	1.6
Start-up time	10	10	8	8	8	10	8	6	4
Overall O&M score	20	17.6	12.8	17.2	16	18	15.2	10.8	8
Other									
Aesthetics	5	4	4	4	3	3	3	2	4
Overall Other Score	5	4	4	4	3	3	3	2	4
Total Score	100	83	70.2	83.2	80	78.2	52.8	43.4	60.6

FIGURE 22 DESIGN MATRIX

Score	1	2	3	4	5
Removal Rate	<60%	<80%	<90%	<98.5%	>98.5%
Operating Costs	>$500/year	<$400/year	<$300/year	<$200/year	<$100/year
Startup Costs	>$25,000	$10,000-$25,000	$5000-$10,000	$2500-$5000	<$2500
Handles Variable Flows	Does Not	With Difficulty	Needs Adjustment Time	Pretty Well	Very Well
Handles Variable Concentrations	Does Not	With Difficulty	Needs Adjustment Time	Pretty Well	Very Well
Retention Time	>36 hours	<24 hour	<12 hour	<8 hour	<4 hour
Complexity	Very Complex	Complex	Medium Complexity	Simple	Very Simple
Maturity	Lacking Data	<10 years old	<20 years old	>20 years old without winery application	>20 years old with winery application
Skill Required	Part-time Engineer	Full time Wastewater Technician	Part time Wastewater Technician	Training needed, infrequent adjustment	Minimal training needed
Material Availability	Material is available by one manufacturer in another country, manufactured <6months	<3 months	Greater than 5 manufacturers <2 months	<1 month	Multiple manufacturers locally sourced, manufactured <1 week
Start-up Time	>300 days	<200 days	<100 days	<50 days	<10 days
Aesthetics	Intolerable odor	Strong odor	Distinct Odor	Weak odor	No odor

FIGURE 23 DESIGN MATRIX SCORING RUBRIC

A2. Lab Results

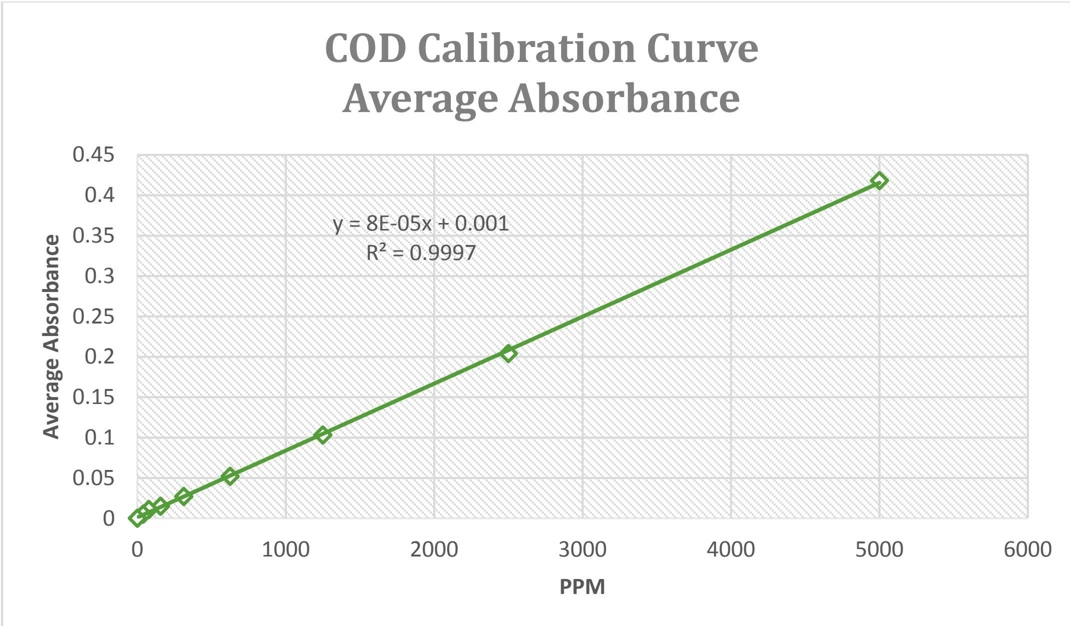

FIGURE 24 COD CALIBRATION CURVE

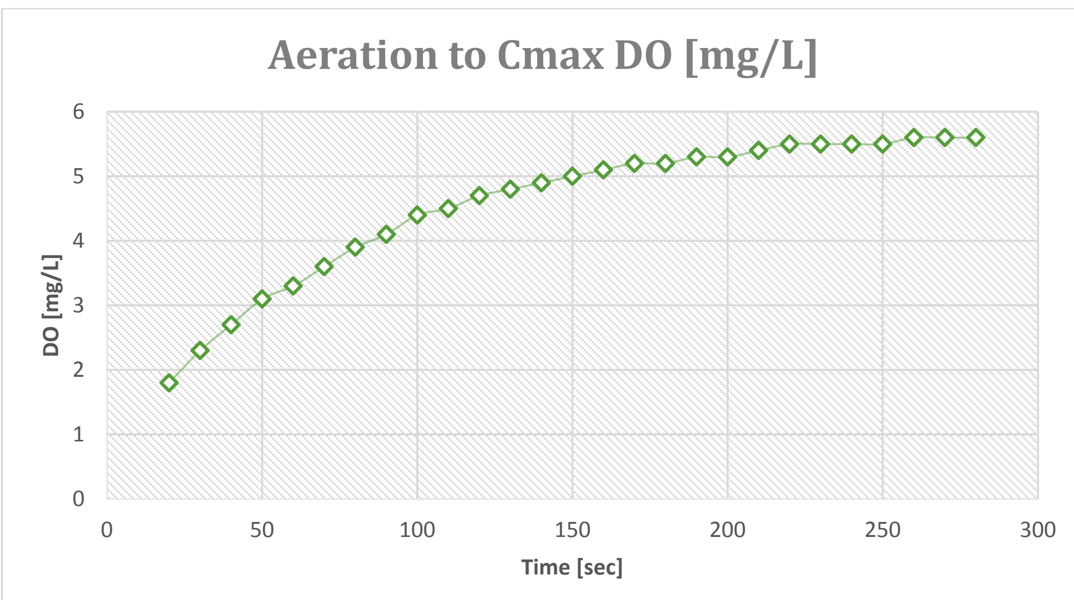

FIGURE 25 AERATION TO MAX

A3. Design Calculation Equations

Hydraulic Design:

$$HRT = V/Q$$

Experimental Oxygen Consumption Rate:

$$\frac{dC}{dt} = -K_d * C$$

$$\ln\left(\frac{C}{C_o}\right) = -K_d * t$$

$$K_d = -\frac{\ln\left({}^{C}/_{C_o}\right)}{t}$$

Modeled:

$$\frac{dC}{dt} = K_{La}\left(C^{*} - C\right) - K_d * C$$

$$\frac{dC}{dt} + C(K_{La} + K_d) = K_{La} * C^{*}$$

$$\frac{dCe^{(K_{La}+K_D)t}}{dt} + C(K_{La} + K_d)e^{(K_{La}+K_d)t} = K_{La}C^{*}e^{(K_{La}+K_d)t}$$

$$Const = C_o - \frac{K_{La}C^{*}}{K_{La} + K_d}$$

$$Ce^{(K_{La}+K_d)t} = \frac{K_{La}C^{*}\, e^{(K_{La}+K_d)t}}{K_{La} + K_d} + C_o - \frac{K_{La}C^{*}}{K_{La} + K_d}$$

$$C = \frac{K_{La}C^{*}}{K_{La} + K_d} + \left(C_o - \frac{K_{La}C^{*}}{K_{La} + K_d}\right) * e^{-(K_{La}+K_d)\,t}$$

A4. Aeration Requirements:

BOD Removal for CSTRs in Series:

$$BOD = \frac{BOD_{in}}{(1 + K_d * \tau)^n}$$

BOD_{in} = The BOD concentration of the influent (mg/L)

τ = Hydraulic retention time (days)

k = BOD removal rate (d^{-1})

Oxygen Requirements (based on BOD loadings):

$$AOR_{max} = \left(BOD_{max} - BOD_{req}\right) * \left[Q_{max} * 3.79\frac{L}{gal} * \frac{1}{10^6}\frac{kg}{mg} * 2.2\frac{lb}{kg}\right] * \left[\frac{1}{1440}\frac{day}{min}\right] * C_{fine}$$

AOR_{max} = Actual Oxygen Required with maximum BOD loading

C_{fine} = 1.10, adjustment factor for fine bubble diffuser system (Red Valve, 2002)

$\frac{AOR}{SOR} = 0.45$, standard adjustment for clean water SOTE to wastewater OTE (Red Valve, 2002)

$$SCFM = \frac{AOR_{max}}{[AOR/SOR]} * [SOTE * d] * 0.0173\left[\frac{lbO2}{SCFM}\right] * FS$$

FS = 1.2 , factor of safety

A5. Nutrient Supplementation

$$Nitrogen\ Needed - Using\ 97{,}400\frac{gal}{yr}\ annual\ flow, BOD_{Vintage} = 1560\frac{mg}{L}$$

$$BOD_{Non-Vintage} = 1440\frac{mg}{L}, \qquad TKN_{Vintage} = 2.4\frac{mg}{L}, TKN_{Non-Vintage} = 4.7\frac{mg}{L}$$

$$Vintage\ flow = .63 * 97400 = 61{,}362\ gal \qquad Non - vintage = .37 * 97400 = 36{,}038\ gal$$

$$Nitrogen\ Needed\ Vintage = 1560\frac{mg}{L} * .05 - TKN_{Vintage} = \frac{78mg}{L} - \frac{2.4mg}{L} = 75.6\frac{mg}{L}$$

$$Phosphorous\ Needed\ Vintage = \frac{1560mg}{L} * .01 - P_{effluent} = 14.4\frac{mg}{L}$$

$$Total\ N\ Vintage = 61362\ gal * 3.78\frac{L}{gal} * .0000756\ kg = 17.5\ kg$$

$$Total\ P\ Vintage = 61362\ gal * 3.78\frac{L}{gal} * .0000144\ kg = 3.34\ kg$$

$$Nitrogen\ Needed\ Non - Vintage = 1440 * .05 - 4.7 = 67.3\frac{mg}{L}$$

$$Phosphorous\ Needed\ Non - Vintage = 1440 * .01 - P_{effluent} = 1440 * .01 - 1 = 13.4\frac{mg}{L}$$

$$Total\ N\ Non - Vintage = 36038\ gal * 3.78\frac{L}{gal} * .0000673\ kg = 9.17\ kg$$

$$Total\ P\ Non - Vintage = 36038\ gal * 3.78\frac{L}{gal} * .0000134\ kg = 1.83\ kg$$

$$\boldsymbol{Total\ N\ Supplementation = 17.5\ kg + 9.17\ kg = 27kg}$$

$$\boldsymbol{Total\ P\ Supplementation = 3.34\ kg + 1.83\ kg = 5.2\ kg}$$

A6. Testing Timeline and Complications

2/8/2017	• Innoculaction • 10 g Sucrose/ 5gal
2/11/17	• Flow rate needs to be doubled • Nutrient load double than needed (1000ppm)*
2/14/2017	• Reduced air flow by adding a splitter to reduce shearing on microbes • Start of testing @ 11 AM • 500 ppm Synthetic Waste*
2/15/2017	• 1000 ppm Synthethic waste started*
2/16/2017	• Flow rate 14 gal/ 12 hours
2/21/2017	• 2000 ppm Synthethic waste started
2/22/2017	• Flow rate 5.3 gal/ 24 hours
2/23/2017	• 5000 ppm Synthethic waste started • Second innoculation**
2/27/2017	• Emptied sedimentation tank
3/1/2017	• Pump Failure - 9.8 gallons remain • Changed arduino flow rate • Pump fixed
3/2/2017	• Started filtering samples for COD sampling
3/4/2017	• 1 unfiltered sample • 2 filtered samples
3/5/2017	• Pump Failure
3/6/2017	• Flow rate 4 gallons / 4 hours
3/7/2017	• Flow rate fixed to 8 gpd • Switched from Sucrose to Glucose
3/10/2017	• Last day of testing

Figure 26 Testing Timeline

	LEGEND
*	Recycled Sludge
**	Second batch of inoculum Obtained from other team's prototypes.

Figure 27 Testing Timeline Legend