

Article

Treating Low-Concentration Methane Emissions via a Methanotroph-Based Biotrickling Filter: Techno-Economic and Life Cycle Assessment

Waaseyaaban-nooji'iwe Landgren ¹, Robert M. Handler ^{2,*} , David R. Shonnard ² and Mary E. Lidstrom ³

¹ Department of Civil, Environmental, and Geospatial Engineering, Michigan Technological University, 1400 Townsend Drive, Houghton, MI 49931, USA; waase.landgren@gmail.com

² Department of Chemical Engineering, Michigan Technological University, 1400 Townsend Drive, Houghton, MI 49931, USA; drshonna@mtu.edu

³ Departments of Chemical Engineering and Microbiology, University of Washington, 1410 NE Campus Parkway, Seattle, WA 98195, USA; lidstrom@uw.edu

* Correspondence: rhandler@mtu.edu

Abstract

Methane, a greenhouse gas which has a global warming potential 80 times greater than carbon dioxide on a 20-year time scale, greatly contributes to global warming. Removing 1 Gt of atmospheric methane by 2050 would limit global temperature increase from reaching 1.5 °C. Currently, biotrickling filter systems for removing atmospheric methane via methanotrophs exist, but not for very low methane concentrations (<1 v%). Recent work at the University of Washington to isolate and improve a microbial strain which thrives at 500 ppmv CH₄ has removed one obstacle in making this technology feasible. In this study, techno-economic and environmental life cycle assessment analyses conducted on this process have assessed its economic feasibility, greenhouse gas reduction potential, and possible areas of improvement. Study results show that at 500 ppmv CH₄, this process could remove atmospheric methane at a cost of USD 3992–5224/tCH₄. The best-performing case also produces annual net reductions in warming potential by 276–311 tCO₂e/120 m³ process unit deployed. Many opportunities exist to improve the outcomes of the baseline analysis even further, especially related to reducing the transport distance of media and harvested biomass.

Keywords: biofiltration; life cycle assessment (LCA); economic feasibility; methane capture; methanotroph



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1. Introduction

As of 2022, methane is the second largest contributor to global greenhouse gas emissions at 11.1% of total emissions, ahead of nitrous oxide (6.1%), and behind carbon dioxide (79.7%). Since the Industrial Revolution, methane is estimated to have contributed about a third of net global warming. On a 20-year time scale, methane has proven to have a warming impact 86 times greater than carbon dioxide. Projections have estimated that removing a total of 0.3–1.0 Gt CH₄ by 2050 will slow global warming by 0.22 °C [1].

Technologies exist to remove methane from the environment, but each technology comes with its own set of limitations. Thermal oxidation operates at more than 1400 °F, demanding high energy consumption [2], so this strategy is typically only employed for point sources of high-concentration methane streams. This is a major disadvantage, and

catalytic oxidation was developed as a means of mitigating these requirements [3]. Catalytic oxidation can operate at lower temperatures (~ 1000 °F) by making use of a catalyst to aid the reaction. While still suffering from high energy consumption, catalytic oxidation is also vulnerable to catalyst poisoning. The catalysts used are often made from precious metals and are expensive to replace [4]. An assessment of methane oxidation methods reports that, even with a 99% efficient heat exchanger capturing the thermal energy from catalytic oxidation, the remaining energy demand would result in 4 °C of global warming, defeating the purpose of the application [5]. Photocatalytic oxidation utilizes catalysts that react to light, which aid in oxidizing methane. However, it suffers from slow rates of methane conversion; even if the rooftops of every building on Earth were covered with photocatalytic paint, the annual methane removal would not outpace the rate of increasing emissions [5]. As with other catalytic oxidation methods, photocatalysis also suffers from catalyst poisoning [6].

Biotrickling filtration already makes use of methanotrophs to oxidize methane. However, current methanotroph biofilters are designed for those bacterial strains that thrive in the 1–5 v% CH₄ range [7], and not for low concentrations such as the 500 ppmv (0.05 v%) value used in this work. Previous work by researchers in this space has identified *Methylobacterium buryatense* 5GB1C as a methanotroph strain that is able to consume 500 ppmv CH₄ at greater rates compared to others [8].

As this biotrickling process requires external resource inputs, the environmental and economic impacts of those inputs will also need to be assessed. The inputs include, but are not limited to the electricity, water, and natural gas utility demands, the cellular nutrient medium, and the trucking services used for shipping. This is an important aspect to consider, as unintended adverse environmental impacts must be avoided. It also must be demonstrated that this process is not prohibitively expensive.

Ref. [9] conducted a techno-economic assessment (TEA) study on a variety of methane mitigation technologies. Thermocatalytic oxidation has an estimated removal cost of EUR 10,000/tCH₄ at a concentration of 500 ppmv. Most of the cost is a result of the energy required to heat the air for combustion. Photocatalytic oxidation has an estimated removal cost of EUR 6500/tCH₄ at a concentration of 500 ppmv. The largest cost associated with this method is the capital cost of enabling the process, such as painting an existing rooftop with photocatalytic paint. Biofiltration has an estimated removal cost of EUR 24,000/tCH₄ when heating is necessary, and EUR 13,000/tCH₄ when heating is unneeded, at a concentration of 500 ppmv. This cost per ton is derived using the method of calculating the levelized capture cost defined in [9]. The conclusion is that these removal technologies, in their current states, are unlikely to compete due to a low CH₄ oxidation rate or oxidation being too costly as a direct consequence of the low availability of sufficiently high methane concentrations [9].

This work is part of an overall project which aims to develop a reactor which can remove methane at concentrations of 500 ppmv utilizing the metabolic processes of methanotrophs and additionally determine the feasibility of pilot-phase testing. Work has been performed by project partners [10] to select and develop a suitable strain of bacteria which thrives in this 500 ppmv CH₄ environment and achieves significant methane removal per gram of biomass. A bench-scale bioreactor has also been developed with the purpose of characterizing the methane removal capacity of such a system under varying conditions such as differing packing mediums, operating temperatures, and input air flow rates. These experimental results have been used to develop an integrated modeling framework, which includes process simulation, environmental and economic impact assessments, for bioreactor assessment under several scenarios. The associated techno-economic and environmental life cycle assessments will serve to highlight several key process conditions and climatic conditions and their influence on the overall results using a detailed scenario analysis, and

to highlight the areas for future research and development to increase the environmental and economic benefits of the process.

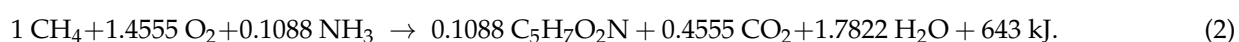
2. Materials and Methods

This CH₄ capture process will utilize packed-bed reactor (PBR) technology in conjunction with methanotrophic bacteria, to produce biotrickling filters capable of removing a significant percentage of CH₄ present in ambient air, which is to be injected into the process. The target removal goal is 1 Mt CH₄/year using the minimum number of reactors needed. A decentralized approach will be taken, and individual reactors will be deployed across numerous sites where ambient air contains at least 500 ppmv CH₄. As a value-added product, bacterial biomass will be harvested routinely from the bioreactors and de-watered on-site. Then, the de-watered biomass will be transported to a more central location to undergo further drying, resulting in the final product. Methanotroph biomass can be further refined to produce supplementary feed for ruminant livestock and salmonids, single-cell protein production, and biofuel production [11–13].

The following biological reaction (1) is required for understanding the yields of biomass, CO₂, water, and energy produced by the consumption of CH₄ within the methanotrophic bacteria being used in this bioreactor system [14]:



The yields of both CO₂ and biomass have been determined experimentally, and so the equation above has been modified to yield approximately the same as the experimental results. Specifically, the reaction stoichiometry has been adjusted most closely to the CO₂ yield, such as to not underestimate the amount of carbon that will be released by the system. It has been determined that the bounds for the reacting O₂ are 1–2 mol, with a typical range of values between 1.4 and 1.8 mol [14]. Additionally, the water and heat generated via this reaction are significant enough to impact the utility usage and thus will not be ignored in subsequent calculations. The modified stoichiometry for the metabolic reaction is listed below in (2), and detailed calculations can be found in Supplementary Materials, Section S1.



The values of constant parameters for this modeling study can be found below in Table 1. Variable parameters that are used to generate various scenarios analyzing the effects of these conditions can be seen in Table 2, alongside the ranges of values used for each parameter.

Table 1. Constant system parameters.

Parameter	Value
Tons CH ₄ removed per year	1,000,000
Working days per year	300
Input air CH ₄ concentration (ppmv)	500
Mass CO ₂ produced/mass CH ₄ consumed	1.25
Mass biomass produced/mass CH ₄ consumed	0.78
kJ produced/kg CH ₄ consumed	40,087
Air density (kg/m ³)	1.293
Water density (kg/ton)	1000

Table 2. Experimental variables and ranges.

Parameter	Range
Reactor volume (m ³)	60–120
Elimination capacity (tCH ₄ /m ³ /yr)	0.025–0.1
Removal efficiency (%)	50–70
Operating temperature (°C)	15–30

Analysis of three locations, Seattle, WA, Knoxville, TN, and Marquette, MI, under 10 different combinations of experimental variable values were performed. These locations were chosen as they represent different climatic classifications that comprise a large portion of the continental U.S. Under the Köppen climate classifications [15], Seattle is located within a warm-summer Mediterranean climate, Knoxville is within a humid subtropical, and Marquette is in a humid continental mild summer, wet all-year region.

A total of 10 unique scenarios have been generated, adjusting one parameter from the baseline conditions of scenario A. These scenarios and their respective variable values can be found below in Table 3. In scenario J, the operating temperature is allowed to fluctuate between 15 and 30 °C, and heating or cooling is only utilized if the reactor temperature is anticipated to move outside of that range.

Table 3. Experimental variable scenarios.

Scenario	A	B	C	D	E	F	G	H	I	J
Reactor volume (m ³)	120	60	90	120	120	120	120	120	120	120
Elimination capacity (tCH ₄ /m ³ /yr)	0.1	0.1	0.1	0.025	0.05	0.1	0.1	0.1	0.1	0.1
Removal efficiency (%)	60	60	60	60	60	50	70	60	60	60
Operating temperature (°C)	20	20	20	20	20	20	20	15	30	15–30

A diagram representing the biomass production process is shown in Figure 1. The process begins with the intake of the surrounding air, driven by a fan through the necessary heater and/or air conditioning unit to add or remove heat. This brings the air temperature within the desired operating range. The air is then injected into the bottom of the reactor and is bubbled through diffusers, where it is allowed to pass up through the BioBall (Aquascape, Chicago, IL, USA) packing medium. Attached to the BioBalls is the biofilm, which strips CH₄ from the air as it passes to the top of the reactor. The air, now saturated with water vapor, is vented off the top of the reactor, released freely back into the surrounding environment. The biomass is allowed to grow until it is harvested at regular intervals, currently at a baseline of 2 weeks, where it is dewatered via a centrifuge on-site, reaching a concentration of approximately 25% solid biomass. Based upon experimental data collected by research colleagues, approximately 80% of the biomass can be dislodged from the packing medium via vortexing; at least 3 iterations are necessary to remove this percentage [10]. It is necessary to leave some of the biofilm present so that the reactor may continue to operate and regrow for the next harvest, so it is not necessary nor desirable to try to harvest 100% of the biomass. The dewatered biomass is then transported via semi-truck to a central location where multiple reactors' harvests can be processed via a drum dryer. The dryer achieves product concentrations of approximately 85% solid biomass, which is then transported to be sold as the final product of the process.

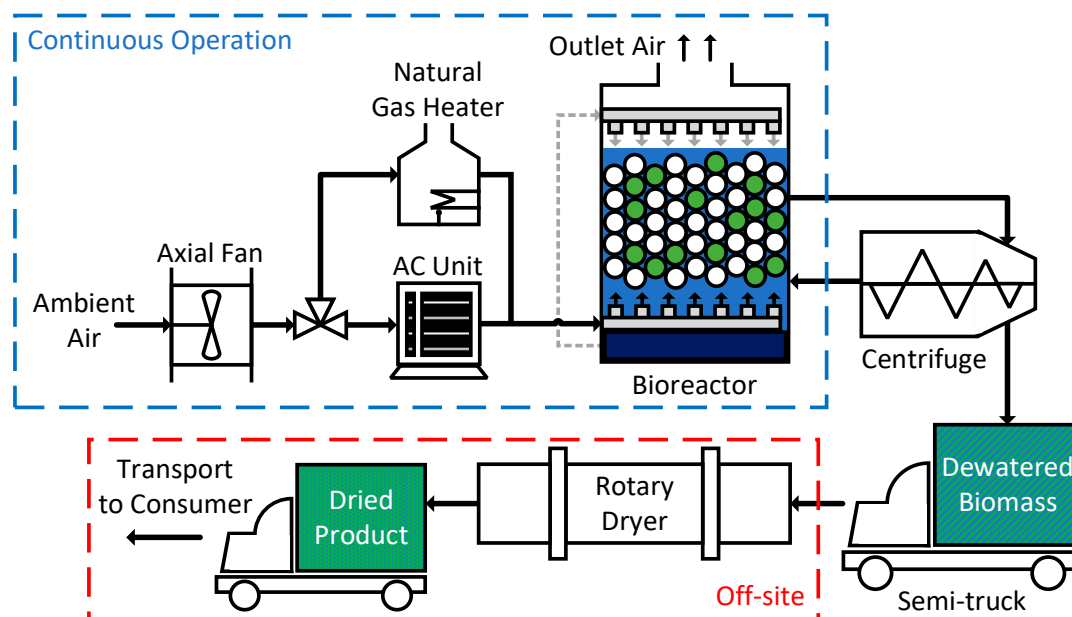


Figure 1. Biomass production process flow diagram.

Table 4 contains information for single-reactor inputs, outputs, and residence times, along with the minimum total reactors required per scenario to achieve the annual 1 Mt CH₄ removal goal. The detailed process of calculating all of these operational characteristics can be found in the Supplementary Materials Document, Sections S1 and S2.

Table 4. Scenario reactor requirements.

Scenario	A	B	C	D	E
Reactors to meet annual removal	83,334	111,112	166,667	166,667	333,334
Air in (kg/hr)	10,950	8212	5475	5475	2737
CH ₄ in (kg/hr)	2.78	2.08	1.39	1.39	0.69
CH ₄ removed (kg/hr)	1.67	1.25	0.83	0.83	0.42
CO ₂ out (kg/hr)	2.08	1.56	1.04	1.04	0.52
Biomass out (kg/hr)	1.30	0.98	0.65	0.65	0.33
Air in (CFM)	4984	3738	2492	2492	1246
Residence time (min)	0.85	0.85	0.85	1.70	3.40
Scenario	F	G	H	I	J
Reactors to meet annual removal	83,334	83,334	83,334	83,334	83,334
Air in (kg/hr)	13,140	9385	10,950	10,950	10,950
CH ₄ in (kg/hr)	3.33	2.38	2.78	2.78	2.78
CH ₄ removed (kg/hr)	1.67	1.67	1.67	1.67	1.67
CO ₂ out (kg/hr)	2.08	2.08	2.08	2.08	2.08
Biomass out (kg/hr)	1.30	1.30	1.30	1.30	1.30
Air in (CFM)	5981	4272	4984	4984	4984
Residence time (min)	0.71	0.99	0.85	0.85	0.85

A notable takeaway from Table 4 is the low gas residence times in the reactor. Recent research in this area has suggested that reactor times of 7–10 min provide optimal methane removal [10], and the residence times found here are between 0.7 and 3.4 min. This could limit the methane elimination capacity of each reactor, which could require building larger reactors, or fundamentally altering the design or operations of the reactor to reduce the residence time requirement. As is, the modeling work presented here relies on the methane elimination capacities presented in Table 4, and should be interpreted as a theoretical

maximum for the given reactor configuration. Future research and development in this area to overcome this issue will be addressed below.

Ambient air is driven by a fan through a natural gas-powered heater and/or an air conditioning unit. Whether both the heater and air conditioning unit are needed depends on the operating temperature of the reactor, and the climate of the location where the reactor is operating. Ambient air is assumed to be at the average monthly temperature for the location.

The heat flux across the reactor vessel accounts for the heat generated by the bacterial metabolic reaction, the heat added/removed via the airstream-reactor temperature difference, and the heat lost via the evaporation of water from the reactor. The equation for determining the net heat flux in a particular month for one of the locations is shown in (3).

$$\dot{Q}_{\text{net}}^{x,y} = \dot{Q}_{\text{rxn}} \left(\frac{\text{kJ}}{\text{hr}} \right) + \dot{m}_{\text{in,air}} \left(\frac{\text{kg}}{\text{hr}} \right) * C_{p,\text{air}} \left(\frac{\text{kJ}}{\text{kg} \cdot ^\circ\text{C}} \right) * (T_{\text{avg}} - T_{\text{op}}) [^\circ\text{C}] - \frac{\Delta H_{\text{vap,H}_2\text{O}} \left(\frac{\text{kJ}}{\text{mol}} \right)}{\text{MW}_{\text{H}_2\text{O}} \left(\frac{\text{g}}{\text{mol}} \right)} * \dot{V}_{\text{air}} \left(\frac{\text{m}^3}{\text{hr}} \right) * r_{\text{evap}}^{x,y} \left(\frac{\text{g H}_2\text{O}}{\text{m}^3 \text{ air}} \right). \quad (3)$$

The heating/cooling demands for scenario J are calculated differently than the other scenarios, as scenario J operates between a range of temperature (15–30 °C) as opposed to a single temperature. Each month, scenario J assumes each reactor starts at the average of the temperature range (22.5 °C).

Net heat flux data from scenario H data can be used to show the necessary heating requirements if the reactor reaches 15 °C. Similarly, scenario I data can be used to show the cooling requirements when the reactor reaches 30 °C. In instances where scenario H (15 °C) has a positive heat flow and scenario I (30 °C) has a negative heat flow, there must be an operating temperature value between the two temperatures where the reactor will reach equilibrium, and the net heat flow will be zero. In these instances, it can be assumed that operating a heating or cooling unit will be unnecessary. For example, during April in Seattle, scenario H shows that if the reactor began at 15 °C, it would begin gaining heat; in scenario I, beginning at 30 °C, the reactor would start losing heat. Scenario J also shows that starting from the average temperature of 22.5 °C, it would still decrease in temperature. Between the operating temperatures of 15–22.5 °C, the rate of heat flow changes from positive to negative, indicating there must be a temperature where the heat flow is zero. The Solver function in Excel was used in conjunction with (3) to determine what operating temperature between 15 and 30 °C would result in a net flow of zero. Table 5 presents examples of instances where this equilibrium occurs.

Table 5. Thermal equilibrium examples, Knoxville data.

		May	June	July	August
Net heat flow (kJ/hr)	Scenario H	30,899	80,373	120,229	117,208
	Scenario I	−135,317	−85,843	−45,987	−49,008
	Scenario J	−52,210	−2735	37,121	34,100
Equilibrium temperature (°C)		17.788	22.253	25.850	25.577

AC units are chosen for each scenario for each location; units provide a maximum number of tons of cooling (equal to 12,000 BTU/h), but a suitable unit in this case is not chosen based upon the tons of cooling it can supply but is instead chosen based upon the amount of CFM of air it is capable of processing. This amount is determined using a rule of thumb used by HVAC technicians and is equal to approximately 412.5 CFM per ton of cooling [16]. This ensures that the air receives the proper amount of cooling desired. The natural gas heaters for each scenario and location are determined similarly to the AC units

but are now sized based upon whatever month demands the greatest number of BTUs per location. Additional details on all equipment sizing can be found in Supplementary Materials, Section S3.

Capital expenses associated with this part of the process are the axial fan, AC unit, and natural gas heater. Electricity consumption by the three pieces of equipment, as well as natural gas use by the heater comprise the operating expenses for this section.

The biotrickling filter reactor modeled here is multi-phase. This consists of the temperature-adjusted air, which is injected below the BioBall packing media. The air passes through diffusers which adjust the airflow into fine bubbles that rise through the column. These bubbles are assumed to spread evenly across the cross-sectional area of the reactor. Aeration is accomplished with minimal compression required due to the high air void spaces (>70%) that are typically seen in packed-bed reactors of this type [17,18].

The liquid phase consists of the thin water film that covers the BioBalls, which is generated via the recirculating spray, which is ejected downward from above the BioBalls. As the water evaporates, it is renewed constantly via the spray. Due to the low flow rate of water required to be recirculated, the pump that would be needed was not considered in the capital costs of this model.

BioBalls are used to fill the column and are not expected to need replacement over the course of the reactor lifetime. The amount of BioBalls required varies only with reactor volume. The solid phase consists of the BioBalls themselves, on which the methanotrophs grow, forming the biofilm. The CH_4 dissolves into the liquid phase, which contacts the methanotrophs that are then able to metabolize it. Once the air reaches the exit at the top of the reactor, it is assumed to be saturated with water vapor. This has ramifications for all the utilities: water, natural gas, and electricity. A non-negligible amount of water is evaporated and must be refreshed into the system regularly, and the cooling effect from the evaporation increases the amount of natural gas used required to keep the reactor temperature high enough. The cooling effect also reduces the amount of necessary AC use, which keeps a significant source of electricity consumption down. The air exits the top of the reactor freely and returns to the surrounding environment. To effectively supply the air to the reactor vessel, it is necessary to equip the bottom of the column with fine bubble diffusers through which the airstream is injected.

Capital costs associated with this section of the process are the reactor vessel, the BioBall packing, and the diffusers. Water consumption and cellular medium nutrients are the operating expenses for this section. Table 6 displays the attributes of reactor equipment for each scenario, such as cost, dimensions, and quantity of equipment pieces.

Table 6. Reactor equipment specifications.

Scenario	A	B	C	D	E
Reactor volume (m^3)	120	90	60	120	120
Air flow (CFM)	4984	3738	2492	2492	1246
Container(s) used	2 × 40 ft	10 ft + 40 ft HC	40 ft	2 × 40 ft	2 × 40 ft
Container cost	USD 6100.00	USD 4935.00	USD 3050.00	USD 6100.00	USD 6100.00
Number of diffusers	250	187	125	125	63
Diffuser cost	USD 15,410.00	USD 11,526.68	USD 7705.00	USD 7705.00	USD 3883.32
Scenario	F	G	H	I	J
Reactor volume (m^3)	120	120	120	120	120
Air flow (CFM)	5981	4272	4984	4984	4984
Container(s) used	2 × 40 ft	2 × 40 ft	2 × 40 ft	2 × 40 ft	2 × 40 ft
Container cost	USD 6100.00	USD 6100.00	USD 6100.00	USD 6100.00	USD 6100.00
Number of diffusers	300	214	250	250	250
Diffuser cost	USD 18,492.00	USD 13,190.96	USD 15,410.00	USD 15,410.00	USD 15,410.00

NMS2 cellular medium is used to fill each reactor, and it comprises several compounds, which are shown along with their respective concentrations in Table 7. KNO_3 , KH_2PO_4 , and $\text{NaHPO}_4 \cdot \text{H}_2\text{O}$ are currently unique in that although they are present in the medium initially, they are not replenished over time; other chemical products are used as alternative sources of elemental nitrogen and phosphorus, which the initial medium compounds provide. Urea and organic fertilizer are being used as more cost-effective alternatives to provide nitrogen and phosphorus to the bacteria, respectively. The necessary quantity of urea and fertilizer are determined as a function of the amount of biomass produced, rather than just as the ratio of the compound present in the medium recipe. Additional details can be found in Supplementary Materials, Section S4.

Table 7. Nutrient medium concentrations.

Compound	g/m ³
NaCl	750
$\text{MgSO}_4 \cdot 7 \text{H}_2\text{O}$	20
$\text{CaCl}_2 \cdot 2 \text{H}_2\text{O}$	1.4
KNO_3	1000
KH_2PO_4	108.8
$\text{NaHPO}_4 \cdot \text{H}_2\text{O}$	214.6
NaHCO_3	588
Na_2CO_3	314
$\text{FeSO}_4 \cdot 7 \text{H}_2\text{O}$	0.4
$\text{CuCl}_2 \cdot 2 \text{H}_2\text{O}$	0

Nutrients are replenished at every reactor harvest, and based upon laboratory experiments conducted by research colleagues [10], adding 10% of the initial nutrient mass every 2 weeks was sufficient for the bacteria to thrive.

Experiments were conducted by research colleagues to determine a feasible method of dislodging the biofilm from the packing media, and it was found that vortex mixing the packing media (cellulose beads used at the time as opposed to the BioBalls assumed in this model) removed up to 78% of the biofilm in a single treatment, with successive treatments improving removal further. It is assumed that this same method would achieve similar results with BioBalls.

The total amount of biomass accumulated per harvest is not equal to the amount that is harvested, as only approximately 80% of the biomass is able to be liberated from the packing medium. This is acceptable because sufficient biomass must be left in the reactor to allow the regrowth of the bacteria and ensure the continual operation of the system. No capital or operating expenses are associated with this part of the process.

Per harvest, water removed from the solution is assumed to be recycled back into the bioreactor, saving on resource use and utility costs. The dewatered intermediate biomass product is then moved on to the transportation steps. At the known yield of 0.78 g biomass/g CH_4 reacted, approximately 218.4 g of biomass accumulates each week. After collecting 80% of the biomass present at each harvest interval, the percentage of solids (biomass) in the solution increases by 0.14–0.15% per week between harvests. At the baseline assumption of 2 weeks between harvests, the accumulated solids percent is 0.29%. Once processed by the centrifuge, the intermediate biomass product contains 25% solids. Recycled water/medium has been factored into the cost of total monthly water demand of each scenario. The centrifuge is the only capital expense for this process section, along with its electricity use being the only operating expense.

Once the on-site centrifuge has removed as much water as possible from the biomass solution, it is prepared to be shipped via semi-truck. It is assumed that a shipping service

will be hired to transport the intermediate product to a central regional facility, which can further dry the biomass solution to its final, saleable form. The shipping service is assumed to cost USD 2.86 per mile traveled, and that each truck will travel an average distance of 100 miles per shipment [19].

The annual cost of shipping the intermediate product to the drying facility is calculated using the average distance per trip, the trucking rate, and the number of harvests per year. No capital expenses are associated with this section, and the annual shipping cost is the only operating expense. Final transportation of the finished product to the consumer is not considered in both emissions and economic calculations in this study, but rather only the intermediate transportation from reactor to drying location.

A rotary drum dryer will be used to process the biomass solution to its final stage, achieving stable dry biomass (85% solids content) once complete. It is assumed that the rotary drum dryer at each central facility will be able to service multiple shipping truck loads in succession, thus there is no need to equip each reactor site with a dryer. Due to this stage of the process demanding >25 kW, the Marquette, MI electric utility is adjusted to USD 8.17/kW per month plus USD 0.16047/kWh. Additional information on all utility demand calculations and related economic and environmental impacts for each stage of the process can be found in Supplementary Materials, Sections S4, S6, and S13.

The centralized drying locations are assumed to require the rental of commercial space; for the bare minimum facility area required, it is assumed that enough space for five 20 ft semi trailers' worth of product to be stored on-site at any one time is needed. An annual rate of USD 8.43/ft² was assumed for the rental space [20], an area of 159 ft² was used for a single 20 ft trailer based upon the given internal dimensions [21], and the dryer takes up 75 ft² of space as based upon manufacturer specifications [22]. With five trailers and the dryer, the facility has a minimum area of 870 ft². At USD 8.43/ft², the annual facility rent is USD 7334.10. As this facility serves 10 reactors, the facility rent costs can be split over each, giving an annual facility rent of USD 733.41 per reactor.

An average of 10 reactors are assumed to be operating within a 100-mile radius of every necessary facility. Shown in Figure 2 are four examples of regions across the U.S. near our scenario locations where a combination of landfills and dairy feedlots are present, demonstrating the viability of this 100-mile radius assumption. As of 2021, a combined total of 13,000 inactive and active landfills were present in the U.S. [23]. Total cattle farms in 2022 were estimated at 732,123, with approximately 12,000 of those being >1000 head farms [24,25]. Stripper wells (oil and natural gas) were estimated at 759,905 in total for 2021 [26]. In 2025 there are currently 14,800 wastewater treatment sites in the U.S. [27]. In most parts of the country, more than 10 potential bioreactor sites would presumably exist within this baseline 100-mile shipping radius being currently assumed in this study.

Like the assumption that the drum dryer can serve multiple reactors' intermediate product shipments, it is assumed that personnel that operate the reactors and drying facilities are able to serve multiple locations. A process engineer is not required to be hands-on with any system to the degree that an operator needs to and is likely only to inspect each location briefly outside of special cases. As a result, it is assumed that the engineer can serve double the number of locations an operator is able to. Labor costs do not change per scenario or location (the US average is used), and the values presented in Table 8 represent the labor expenses after being spread across the appropriate reactors serviced.

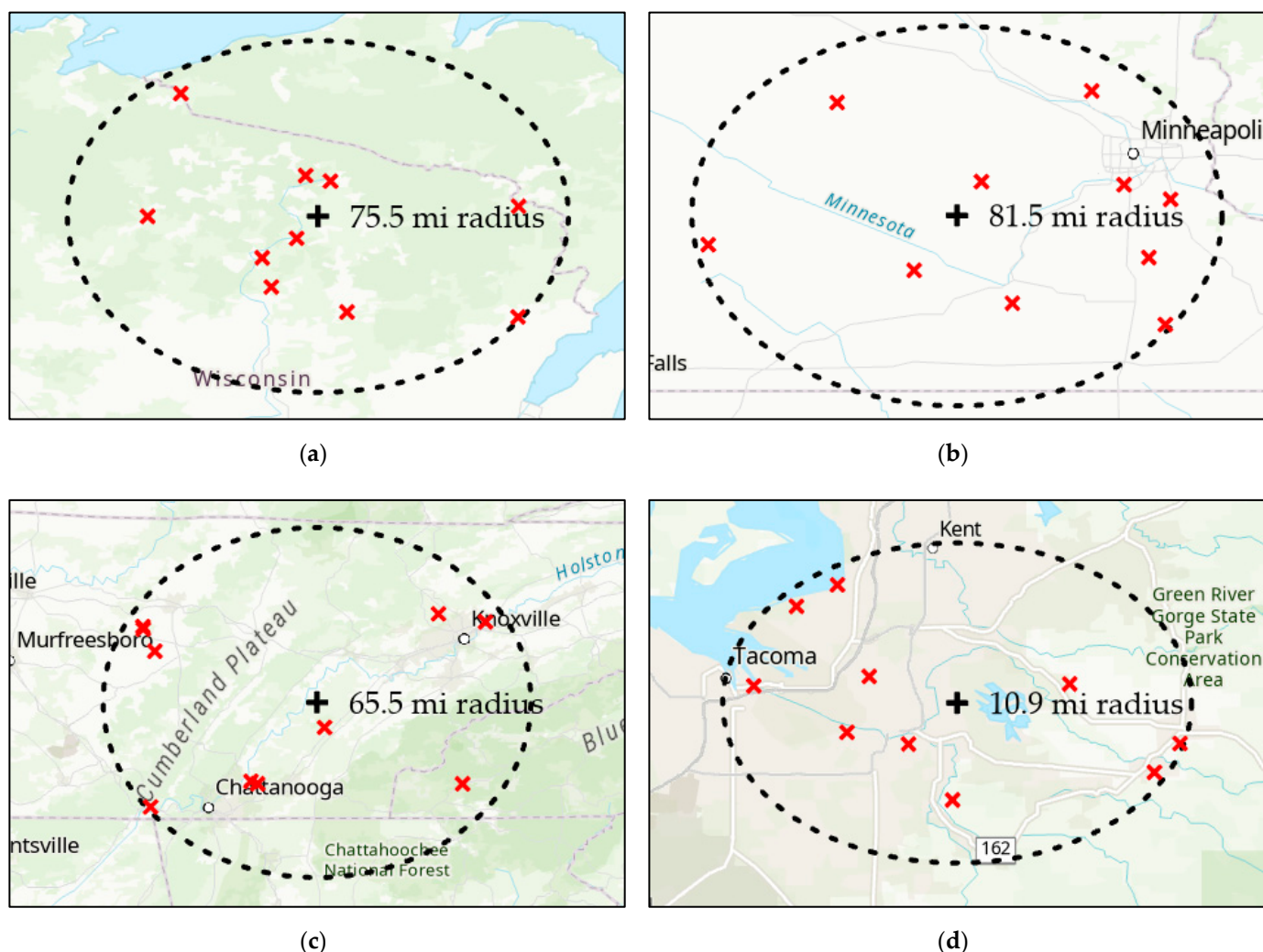


Figure 2. (a) A total of 10 wastewater treatment plants and landfills within approximately 76 miles of Rhinelander, WI; (b) 10 landfills within approximately 82 miles of Glencoe, MN; (c) 10 landfills within approximately 66 miles of Sweetwater, TN; (d) 10 wastewater treatment plants within 11 miles of Sumner, WA.

Table 8. Estimated annual labor costs per reactor.

Position	Reactors Serviced	US Average
Chemical/Process Engineer	20	USD 5605.00
Chemical Plant and System Operator	10	USD 8003.00
Total		USD 13,608.00

The drum dryer is the only capital expense associated with this part of the process, and the operating expenses consist of the facility rent and the electricity consumption by the dryer.

One value-added product generated by the process includes the dried, 85% biomass product in 15% moisture, which can be used for multiple purposes, such as a single-cell protein for consumption by both humans and livestock, undergoing further processing to create biofuels, or serve as fertilizer for crops. For this study, biomass is valued at USD 1600/ton [28], which assumes a 1:1 replacement for dried, stable fishmeal based on the protein content of the animal feed. The biomass product generated by this process offers a suitable replacement for fishmeal, which typically contains 50–60% protein by mass.

Methanotrophs and methanotroph-based single cell protein have been found to contain 59–81% crude protein content [11,29], so the protein content of the product here is assumed to be sufficient for fishmeal replacement. There may be other incidental costs associated with packaging and regulatory approval, but it is assumed that those costs would be minor and roughly equivalent between conventional fishmeal and this new animal food substitute product.

The annual revenue generated by the production of biomass is calculated via (4). The other source of revenue assumes the existence of and access to a market of carbon credits. Since there is no nation-wide market, the value of removing 1 tCO₂e from California's cap-and-trade market is being used to estimate the revenue stream. As of August 2024, that market value was USD 34/tCO₂e, but this price has recently fluctuated between USD 25 and USD 40 [30]. The theoretical revenue from carbon credits is calculated using (5).

$$\text{Annual biomass revenue} = m_{\text{BM,H}} \left(\frac{\text{kg}}{\text{harvest}} \right) * N_{\text{H,a}} \left(\frac{\text{harvests}}{\text{yr}} \right) * \frac{1 \text{ kg}}{1000 \text{ g}} * \frac{\text{USD}1600}{\text{ton BM}}. \quad (4)$$

$$\text{Annual carbon credit revenue} = m_{\text{a,tCO}_2\text{e}} \left(\frac{\text{tCO}_2\text{e}}{\text{yr}} \right) * \frac{\text{USD}34}{\text{tCO}_2\text{e}}. \quad (5)$$

The environmental life cycle assessment studies being conducted in this work have been completed in accordance with ISO [31,32] guidance. The goal of the LCA work was to understand the greenhouse gas emissions associated with the bioreactor process, and the system boundary utilized for this work is consistent with the techno-economic work as described in prior sections and Figure 1. The functional unit of concern is 1 ton of CO₂e emission removal performed by the bioreactor system, to which the environmental impacts are scaled. Additional details are to be found in Sections 2 and 4 of the work, along with Supplementary Materials Sections S5 and S13.

3. Results

The analysis of CAPEX and OPEX assumes the following: the base year is 2024, linear depreciation will be used for CAPEX over a span of 10 years, reactor life is also assumed to be 10 years, and reactors operate 300 days/year (split evenly across all 12 months), 24 h/day.

3.1. Capital Costs

A breakdown of capital costs for all scenarios based on the Washington location is shown below in Table 9. Other locations follow a similar cost structure, with notable differences related to heater or AC sizing based on climate considerations, and can be seen in the Supplementary Materials Document, Section S12. A detailed breakdown of equipment sizing calculations can be seen in Supplementary Materials, Section S3. A 10-year linear depreciation has been applied to the capital costs of long-lasting equipment.

The centrifuge ends up as the single most expensive item per reactor, as one is assumed to be required at every reactor site. Drum dryers are the second most expensive item outright (USD 29,800 assumed cost), but their cost can be spread over the number of reactors it serves (10 is used here as an average assumption), and as a result it is one of the cheapest capital expenses, depending on whether both an AC unit and heater are required. In some scenarios for the WA location, when a cooler operating temperature is required, an AC unit is needed for the system, but in most scenarios with higher temperatures, an AC unit is not needed at this location.

Table 9. Annual CAPEX costs per reactor (WA).

Scenario	A	B	C	D	E
Fan/Blower	USD 99.33	USD 99.33	USD 62.48	USD 62.48	USD 51.90
AC unit	-	-	-	-	-
Heater	USD 185.44	USD 165.78	USD 165.78	USD 165.78	USD 119.79
Reactor vessel	USD 610.00	USD 493.50	USD 305.00	USD 610.00	USD 610.00
Diffusers	USD 1541.00	USD 1152.67	USD 770.50	USD 770.50	USD 388.33
Packing media	USD 1800.00	USD 1350.00	USD 900.00	USD 1800.00	USD 1800.00
Centrifuge	USD 3770.78	USD 3770.78	USD 3770.78	USD 3770.78	USD 3770.78
Drum dryer	USD 298.00	USD 298.00	USD 298.00	USD 298.00	USD 298.00
Total	USD 8304.55	USD 7330.06	USD 6272.54	USD 7477.54	USD 7038.80
Scenario	F	G	H	I	J
Fan/Blower	USD 107.22	USD 99.33	USD 99.33	USD 99.33	USD 99.33
AC unit	-	-	USD 673.70	-	-
Heater	USD 236.24	USD 165.78	USD 165.78	USD 314.85	USD 236.24
Reactor vessel	USD 610.00	USD 610.00	USD 610.00	USD 610.00	USD 610.00
Diffusers	USD 1849.20	USD 1319.10	USD 1541.00	USD 1541.00	USD 1541.00
Packing media	USD 1800.00	USD 1800.00	USD 1800.00	USD 1800.00	USD 1800.00
Centrifuge	USD 3770.78	USD 3770.78	USD 3770.78	USD 3770.78	USD 3770.78
Drum dryer	USD 298.00	USD 298.00	USD 298.00	USD 298.00	USD 298.00
Total	USD 8671.45	USD 8062.99	USD 8958.59	USD 8433.97	USD 8355.36

3.2. Operating Costs

For the ideal scenario (J), the gap between the most and least expensive locations is wider here than it is for CAPEX; USD 8962.61 greater in Marquette than it is for Knoxville. While the CAPEX per location for scenario J differed by only around 0.8%, Knoxville is the cheapest for annual OPEX by a much greater margin. Relative to Knoxville, overall annual OPEX is greater by 9% and 47% for Seattle and Marquette, respectively. Table 10 presents the breakdown of OPEX sources per location, as well as the change in each relative to Knoxville.

Table 10. Change in annual OPEX for operating scenario J relative to Knoxville.

Location	Knoxville, TN	Seattle, WA	Change	Marquette, MI	Change
Total kW demand charges	USD 278.12	-	−USD 278.12	USD 3921.60	USD 3643.48
Fan electricity	USD 1142.16	USD 986.45	−USD 155.72	USD 1275.54	USD 133.38
AC unit electricity	-	-	-	-	-
Heater electricity	USD 144.97	USD 160.20	USD 15.23	USD 471.68	USD 326.71
Dewatering electricity	USD 318.85	USD 275.38	−USD 43.47	USD 356.09	USD 37.24
Drying electricity	USD 372.32	USD 114.28	−USD 258.04	USD 36.95	−USD 335.37
Water	USD 191.62	USD 573.55	USD 381.92	USD 567.31	USD 375.69
Natural gas	USD 1489.87	USD 3575.18	USD 2085.31	USD 6271.34	USD 4781.47
Cell nutrients	USD 1799.42	USD 1799.42	-	USD 1799.42	-
Trucking	USD 6006.00	USD 6006.00	-	USD 6006.00	-
Drying facility rent	USD 7334.10	USD 7334.10	-	USD 7334.10	-
Total	USD 19,077.44	USD 20,824.55	USD 1747.11	USD 28,040.05	USD 8962.61

The 94% of the USD 8962.61 increase from Marquette's OPEX relative to Knoxville can be attributed to two OPEX categories: power demand charges and natural gas consumption. Marquette's power demand rate of USD 8.17/kW/mo for operations > 25 kW is much greater versus Knoxville's USD 0.50/kW/mo rate, adding USD 3643.48 to annual OPEX.

Marquette's more extreme winter climate demands an extra 387,000 BTU/yr over Knoxville. In addition, with a natural gas rate of $\text{USD } 8.51 \times 10^{-6}/\text{kJ}$, almost exactly twice that of Knoxville's rate of $\text{USD } 4.25 \times 10^{-6}/\text{kJ}$, adds USD 4781.47.

For Seattle, essentially all its increase in OPEX relative to Knoxville comes from its greater natural gas consumption, which adds USD 2085.31/year. Water is also more expensive in Seattle, adding another USD 573. This increase is offset slightly by Seattle's cheaper electricity and lack of power demand charges, reducing the total OPEX of electricity-consuming sources by USD 691/year.

The changes from the base scenario are as follows:

- Scenario B: Reactor volume reduced from 120 m³ to 90 m³;
- Scenario D: Elimination capacity reduced from 0.1 to 0.05;
- Scenario F: Removal efficiency reduced from 60% to 50%;
- Scenario H: Operating temperature reduced from 20 °C to 15 °C;
- Scenario J: Operating temperature changed from static 20 °C to a range of 15–30 °C.

The three most influential parameters for optimizing process costs are (ordered most to least influential) removal capacity, reactor volume, and operating within a range of temperatures as opposed to a static one. Summaries of the changes in total scenario CAPEX and OPEX can be found in Table 11.

Table 11. Total OPEX, CAPEX with percent change, Seattle WA.

Scenario	A	B	D	F	H	J
Total reactors	83,334	111,112	166,667	83,334	83,334	83,334
Total OPEX (billions USD)	2.077	2.475	3.230	2.315	2.188	1.735
Total CAPEX (billions USD)	0.692	0.814	1.246	0.723	0.747	0.696
Change (OPEX)	-	19%	56%	11%	5%	−16%
Change (CAPEX)	-	18%	80%	4%	8%	1%

3.3. Environmental Impacts

The software SimaPro (version 9.3) was used to assemble the life cycle inventory (LCI) for the process, as well as conduct the life cycle impact analysis. LCI data was taken from the ecoinvent database [33], and Global Warming Potential impacts were quantified using the IPCC 2021 GWP 100a method. Emissions generated by the system were assumed to result from the following: electricity and natural gas usage, liquid media bioreactor inputs, and the use of trucks for shipping. Additional details related to the emission factors associated with key input items are presented in the Supplementary Materials, Sections S5 and S13. Table 12 contains the breakdown of key utility inputs to the bioreactor process, and key sources of emissions to the process, along with annual net emissions, expressed in tons CO₂e, for each operating scenario modeled at the WA location. Additional results related to other locations can be found in the Supplementary Materials, Section S13, but the results here are illustrative of the general trends. Utility emissions are significant, especially processing heating for biomass drying and temperature control within the bioreactor. The impacts of reactor media nutrients are also significant in each scenario. However, the net benefit associated with reactor operations, oxidizing methane gas to carbon dioxide as part of the metabolic process of the microbes involved, more than makes up for the emissions associated with any part of the process. Net negative emissions are reported for each bioreactor operating scenario reported.

Table 12. Annual emission sources breakdown (WA location).

Scenario	A	B	C	D	E
kWh per year	11,984	10,975	4661	5770	11,916
Nat. gas MJ per year	692,240	519,180	346,121	346,121	173,063
Electricity tCO ₂ e/yr	1.45	1.33	0.56	0.70	1.44
Nat. gas tCO ₂ e/yr	51.92	38.94	25.96	25.96	12.98
tCO ₂ e emitted by rxn	15.00	11.25	7.50	7.50	3.75
tCO ₂ e nutrient production	7.62	5.72	3.81	3.98	2.16
tCO ₂ e urea decomposition	0.25	0.18	0.12	0.12	0.06
Trucking tCO ₂ e/yr	0.43	0.32	0.21	0.21	0.11
tCO ₂ e via CH ₄ conversion	−360.00	−270.00	−180.00	−180.00	−90.00
Net tCO ₂ e emissions	−283.33	−212.26	−141.83	−141.53	−69.50
Scenario	F	G	H	I	J
kWh per year	12,879	11,760	55,153	15,116	12,380
Nat. gas MJ per year	926,896	524,628	331,280	1490,072	320,223
Electricity tCO ₂ e/yr	1.56	1.42	6.67	1.83	1.50
Nat. gas tCO ₂ e/yr	69.52	39.35	24.85	111.76	24.02
tCO ₂ e emitted by rxn	15.00	15.00	15.00	15.00	15.00
tCO ₂ e nutrient production	7.62	7.62	7.62	7.62	7.62
tCO ₂ e urea decomposition	0.25	0.25	0.25	0.25	0.25
Trucking tCO ₂ e/yr	0.43	0.43	0.43	0.43	0.43
tCO ₂ e via CH ₄ conversion	−360.00	−360.00	−360.00	−360.00	−360.00
Net tCO ₂ e emissions	−265.63	−295.93	−305.19	−223.12	−311.19

3.4. Process Economic Feasibility

An important metric for understanding the economic feasibility of this work is the removal cost per tCO₂e and per tCH₄. For this work, the calculation of removal cost per tCO₂e is achieved using (6), which uses the total annual expenses (capital depreciation, operating expenses, labor), the annual revenue from biomass, and the annual net tCO₂e emitted/removed. To find the cost of removal per tCH₄, the value resulting from (6) is converted to tCH₄ assuming that 1 tCH₄ is equal to 30 tCO₂e on a 100-year global warming potential basis. This operation is shown in (7).

$$r_{\text{CO}_2\text{e}}^r = \frac{\text{Exp}_a(\text{USD}) - \text{Rev}_{\text{BM}}(\text{USD})}{m_{a,\text{tCO}_2\text{e}} \left(\frac{\text{tCO}_2\text{e}}{\text{yr}} \right)} \quad (6)$$

$$r_{\text{CH}_4}^r = r_{\text{CO}_2\text{e}}^r \left(\frac{\text{USD}}{\text{tCO}_2\text{e}} \right) * \frac{30 \text{ tCO}_2\text{e}}{1 \text{ tCH}_4} \quad (7)$$

The estimated removal costs of both CO₂e and CH₄ for each scenario in this work are summarized in Tables 13 and 14, respectively. In the ideal scenario for the Knoxville location, the cost of oxidizing methane is USD 2869.63/tCH₄, or expressed in terms of CO₂e, USD 95.65/tCO₂e. The removal cost per tCO₂e can be interpreted as the selling price for carbon credits which would result in an economic break-even point per reactor. Heating is required in all 10 scenarios presented, so relative to the literature estimation [9] of CO₂e and CH₄ removal costs, the ideal scenario shows the possibility of achieving costs that are just 8.61% of the literature values.

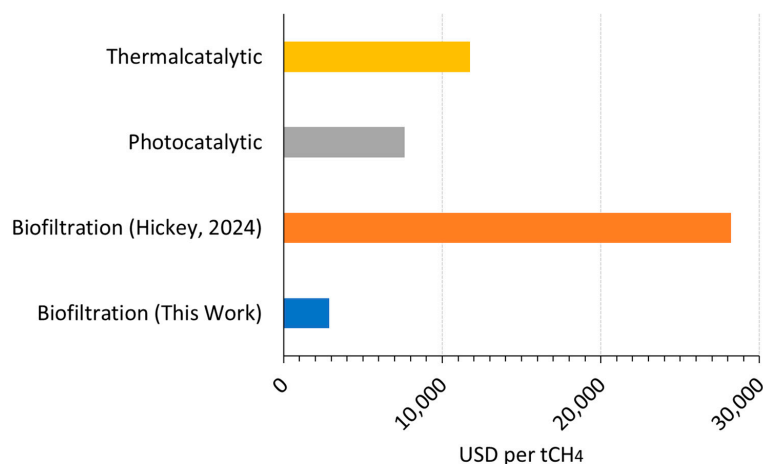
Shown in Figure 3 are the approximate removal costs per tCH₄ for methods analyzed in [9], alongside the ideal removal cost for the process designed here, which occurs in scenario J in Knoxville. All estimated costs are at a concentration of 500 ppmv CH₄. The removal cost for biofiltration in prior work [9] is an order of magnitude greater than the cost associated with this process. In [9] the assumption is made that only approximately 5 tCH₄ can be oxidized by a standard bioreactor (~120 m³) per year at 500 ppmv, which is less than half of the assumed removal per reactor in the ideal scenario presented in this work.

Table 13. Removal cost per tCO₂e.

Scenario	A	B	C	D	E
Seattle, WA	USD 123.88	USD 162.11	USD 232.53	USD 244.46	USD 510.34
Knoxville, TN	USD 144.38	USD 175.25	USD 266.49	USD 279.80	USD 604.79
Marquette, MI	USD 165.10	USD 214.07	USD 299.02	USD 313.57	USD 677.76
Scenario	F	G	H	I	J
Seattle, WA	USD 144.24	USD 111.10	USD 121.50	USD 199.56	USD 99.77
Knoxville, TN	USD 163.47	USD 130.81	USD 142.40	USD 144.90	USD 95.65
Marquette, MI	USD 196.47	USD 176.99	USD 188.27	USD 253.64	USD 138.59

Table 14. Removal cost per tCH₄.

Scenario	A	B	C	D	E
Seattle, WA	USD 3716.48	USD 4863.43	USD 6975.89	USD 7333.90	USD 15,310.12
Knoxville, TN	USD 4331.46	USD 5257.51	USD 7994.62	USD 8393.88	USD 18,143.82
Marquette, MI	USD 4953.06	USD 6422.16	USD 8970.65	USD 9406.95	USD 20,332.92
Scenario	F	G	H	I	J
Seattle, WA	USD 4327.17	USD 3332.88	USD 3645.12	USD 5986.85	USD 2993.06
Knoxville, TN	USD 4904.22	USD 3924.35	USD 4271.86	USD 4347.11	USD 2869.63
Marquette, MI	USD 5894.01	USD 5309.63	USD 5648.23	USD 7609.12	USD 4157.74

**Figure 3.** Removal cost comparison for different methane-oxidizing technologies at 500 ppmv CH₄ [9].

In Table 15, the annual profit per reactor for each scenario and each location are summarized. These values account for all the annual expenses (capital depreciation, operating costs, labor) and the total revenue (including the theoretical revenue from carbon credits valued at USD 34/tCO₂e).

Table 15. Annual profit per reactor.

Scenario	A	B	C	D	E
Seattle, WA	–USD 25,466.85	–USD 27,193.60	–USD 28,157.05	–USD 29,785.92	–USD 33,106.50
Knoxville, TN	–USD 29,749.42	–USD 28,904.53	–USD 31,105.35	–USD 32,762.72	–USD 35,664.39
Marquette, MI	–USD 32,877.38	–USD 33,760.82	–USD 33,486.85	–USD 35,165.17	–USD 38,230.20
Scenario	F	G	H	I	J
Seattle, WA	–USD 29,282.44	–USD 22,815.19	–USD 26,705.11	–USD 36,939.80	–USD 20,466.35
Knoxville, TN	–USD 32,751.69	–USD 27,289.87	–USD 30,779.37	–USD 25,927.45	–USD 18,885.18
Marquette, MI	–USD 36,927.20	–USD 36,926.54	–USD 40,108.77	–USD 41,901.84	–USD 28,932.09

4. Discussion

We have shown the potential for low-cost methane removal using a biotrickling filter deployed at sites with relatively low ambient methane concentrations. As described in the Introduction, many such sites exist throughout the U.S., and the deployment of these reactors may be an appealing method of oxidizing difficult to treat area sources of methane. Although none of the scenarios display a net profit at the baseline assumptions used in modeling, Table 13 shows that under operating conditions where the temperature constraints of the reactor are allowed to fluctuate within the range of acceptable temperatures for our candidate microbe, our anticipated GHG emission mitigation costs can be under USD 100 per ton CO₂, which is competitive with other technologies being evaluated in this space. These GHG mitigation costs are dependent on the sale of the dried biomass product as a fishmeal replacement, which we have valued above at a selling price of USD 1600/ton. If the market for this fishmeal replacement changed so that this product could only be sold for USD 800, the effective CO₂ removal credit price would need to increase by USD 20/ton in order to compensate for the lost revenue.

There are a number of potential improvements that may be considered in future investigations, throughout the entire process platform, but they must be evaluated carefully. Due to the high cost of the centrifuge relative to the other capital expenses, it may appear attractive to attempt to centralize the dewatering step alongside the drying step; being able to spread the cost of the centrifuge over 10 reactor sites would reduce its cost to only USD 377.08/reactor. If the centrifuge were centralized, and assuming at worst, the entirety of the harvested solution would need to be shipped away with none of it able to be recycled back to the reactor. An additional 74.08 CCF of water would be lost per harvest for a 120 m³ reactor. At the Seattle water utility rate of USD 7.28/CCF, at 21 harvests/year, this would add USD 6988.31 of water utility expenses annually. This would negate any benefit of trying to save on capital expenses; over the 10-year reactor lifetime, the increased water use would cost USD 69,883.10. Alternatively, even if some of the medium was recycled from the proposed dewatering and drying facility, trucking it back to the reactors may increase the number of shipments per year, along with the trucking expense. It is better to pay for the high capital cost of a centrifuge for each reactor, centralizing only the drying step and avoiding increasing operating costs to undesirable levels. There are several key assumptions that are being made that have a flexible range of realistic values, such as personnel distribution, shipping distance, or the harvesting interval. For example, Table 16 shows the impact of increasing the number of weeks between biomass harvesting. Trucking expenses are always the second or third largest source of OPEX; increasing the time between harvests greatly reduces the number of shipments per year, cutting costs significantly.

Table 16. Harvest interval impacts, Seattle, WA, scenario J.

Harvest Interval (Weeks)	2	3	4
Harvests per year	21	14	10
Annual trucking expenses	USD 6006.00	USD 4004.00	USD 2860.00
Removal cost per tCO ₂ e	USD 99.77	USD 92.39	USD 89.94
Removal cost per tCH ₄	USD 2993.06	USD 2771.68	USD 2698.14

Technologies utilized at several steps of the process are certainly not the most efficient, cutting-edge approaches to accomplish the appropriate unit operations. For example, there are more variations in bioreactors that are not mass-transfer limited as the packed-bed reactor assumed here and would likely offer improved CH₄ removal rates. Centrifuges and drum dryers are mature, well-understood pieces of equipment but suffer from high capital and operating expenses, where advances in other methods of dewatering and drying may

prove to be more cost-effective. The baseline cases evaluated in this work are close to turning a profit, if the assumed biomass selling prices and carbon credit selling prices can be realized. Process economics could be significantly shifted just by adjusting a few system parameters; pursuing any of several possible improvements that could be made to this model would likely result in future profitability. A potential improvement to save on capital expenses per reactor would be to mount the centrifuge on a vehicle that is able to move between reactor sites. This would allow for the dewatering of harvested batches from multiple sites with a single centrifuge, as opposed to having one permanently at every location.

Since the reactor type used in this study has shown to be mass-transfer limited in recent experiments [10], real-world deployments would have to involve larger reactors, which would increase capital costs related to media purchase and reactor sizing. Future studies should focus on increasing methane mass transfer through changing reactor type (e.g., tray-based systems) or bubble shape, or increasing effective gas residence time. Additionally, enhancements to the bacteria used in the work, by increasing the acceptable temperature range, or increasing their metabolic capacity to use methane, would all increase the efficacy of this process. Finally, to build upon this work, a study of a larger-scale system would allow additional experimental data to be collected showing the real performance of a process utilizing the equipment specifications laid out in this work.

5. Conclusions

In this work, a model-based evaluation was conducted to illustrate the potential environmental and economic viability of a biological methane oxidation system that can operate at low ambient methane concentrations, and could be deployed in many locations around the U.S. and the world. The TEA and LCA studies conducted can serve as a baseline for which to compare subsequent work that relies on biological methane oxidation. The application of well-understood wastewater management equipment (disk-stack centrifuges, drum dryers) to biomass cultivation has established a reasonable technical baseline, which may not be the most cost-effective means of methane removal. This study shows that even with conventional technologies, we are quite close to profitability, and future research that evaluates the incorporation of more advanced dewatering and drying systems, or new supply chain logistics, can further improve economic outcomes. Environmentally, the process provides a net reduction in emissions and does not produce CO₂ or other GHGs in quantities that would negate the benefit of capturing CH₄. The potential for meeting the overall annual removal goal of 1 Mt CH₄ is there; the number of suitable locations to deploy this process exceeds the number of units that would be required. As is, this process is estimated to provide removal of CH₄ at a fraction of the cost presented in prior studies; in the ideal case it would require USD 2869.63/tCH₄, or USD 95.65/tCO₂e. At a value of USD 1600/ton of biomass, reducing the processing costs approximately three-fold would allow this process to break even economically without any additional source of revenue, making a potential carbon credit market unnecessary. However, should a carbon credit market arise by which this process could generate revenue regardless of its location, it would offer the potential to offset process expenses significantly. In summary, this direct CH₄ capture process, as modeled, should offer a basis for which to compare future studies that utilize more state-of-the-art technologies.

Supplementary Materials: The following supporting information can be downloaded at <https://www.mdpi.com/article/10.3390/methane4040023/s1>. The Supplementary Materials document associated with this manuscript contains a wealth of information about reactor sizing, material inputs and output calculations, utility requirements, individual cost parameter calculations, emissions factor data, and additional results for specific locations and scenarios that are not able to fit in the main manuscript [34–95].

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Abbreviations

The following abbreviations are used in this manuscript:

ACO	The AC Outlet
ACW	AC Wholesalers
BAU	Business-As-Usual
CAPEX	Capital Expenses
CMUB	City of Marquette Utility Billing
GHG	Greenhouse Gas
GI	Global Industrial
GWP	Global Warming Potential
IPCC	International Panel on Climate Change
KUB	Knoxville Utilities Board
LCA	Life Cycle Analysis
LCI	Life Cycle Inventory
LCIA	Life Cycle Impact Assessment
MPSC	Michigan Public Service Commission
NCEI	National Centers for Environmental Information
NOAA	National Oceanic and Atmospheric Administration
OPEX	Operating Expenses
PBR	Packed-Bed Reactor
PFD	Process Flow Diagram
SCL	Seattle City and Light
SPU	Seattle Public Utility
TEA	Techno-Economic Assessment
TOC	Tons of Cooling
WU	Weather Underground

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