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ANAEROBIC DIGESTION OF LIGNOCELLULOSIC BIOMASS VIA COTREATMENT: A TECHNO-ECONOMIC ANALYSIS

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Abstract

This study evaluates an alternative to pretreatment that represents a biomimetic approach to fermenting recalcitrant cellulosic biomass. This approach is modeled after a biological mechanism that has proven over time to efficiently deconstruct lignocellulosic biomass: the ruminant's digestive system.

In the last century, the world has been paying increasing attention to greenhouse gas (GHG) emissions and climate change, with the agricultural and energy sectors as two of the largest emitters. Lignocellulosic biomass from perennial crops, crop residues, winter crops and manures can reduce or reverse agricultural GHG emissions relative to conventional summer annual crops like maize and soybean. Renewable Natural Gas (RNG) produced by anaerobic digestion (AD) of lignocellulosic biomass can be a sustainable alternative to fossil natural gas to reach renewable energy policy goals. Conventional AD of lignocellulose is usually not cost competitive relative to fossil fuels, largely due to the long-residence times and hence large digester volumes required to convert recalcitrant cellulosic feedstocks. Recent research with pure cultures suggests that mimicking rumination by milling intermittently during fermentation can improve lignocellulose digestibility and has the potential to lower cost by increasing yield and/or by reducing retention time. Our study is motivated by the possibility that this biomimetic strategy, termed cotreatment, can similarly improve AD.

Techno-economic assessment of the process is still needed to assess scale-up viability and potential economic implications of cotreatment assisted AD for renewable natural gas production. The following research intends to assess scale-up viability of cotreatment assisted switchgrass fed anaerobic digestion. Two phases are carried out: technical process modeling and economic analysis. Sensitivity analysis was carried out to study potential impacts of carbohydrate solubilization (due to cotreatment) and scale on minimum fuel selling price ($ GGE^{-1}).

For the process modeling phase, an Aspen Plus model was developed to determine mass and energy flows of each process area. Mass flow results show a potential increase of 27% more biomethane production with cotreatment in comparison to no cotreatment for a fixed ten day residence time. These results served as input parameters for the second phase of economic analysis.

The cost for cotreatment aided, mixed culture, biomass-fed anaerobic digestion systems for biomethane production at a scale of 2000 Mg day^{-1} is \$3.37 GGE^{-1} compared to \$4.23 GGE^{-1} for no cotreatment (2014 USD). Thus, cotreatment decreases the MFSP by \$0.86 at that scale. The carbohydrate solubilization sensitivity analysis estimates a 5 cent reduction in MFSP per 1% increase in solubilization. At a 94.4% carbohydrate solubilization factor and 2000 dry Mg day^{-1} scale, cotreatment aided switchgrass fed AD becomes cost competitive relative to CNG. However, when adding a RIN incentive of \$3.36 GGE^{-1} of CNG to a market fossil CNG price of \$2.09 GGE^{-1}, switchgrass fed AD becomes economically feasible at a scales greater than 230 dry Mg day^{-1} without cotreatment, and at scales greater than 104 dry Mg day^{-1} with cotreatment.

These results indicate that under current prices and reasonable conversion assumptions cotreatment could be a favorable option for production of biomethane from biomass sources. These findings can aid future planning of large-scale anaerobic digesters to reach government renewable energy policy targets, reduce greenhouse gas emissions, and provide a sustainable, cost-efficient bioenergy resource.

Table of Contents

List of Figures

List of Tables

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Chapter 1: Introduction

Bioenergy is energy produced from biorenewable resources (organic materials of recent biological origin, such as energy crops and organic wastes), and different conversion technologies can be used to produce heat, power or transportation fuel (Brown and Brown, 2014).Bioenergy can enhance energy security by providing an alternative to fossil fuels while reducing greenhouse gas (GHG) emissions. A wide range of processes can be used to generate bioenergy; however, the optimum strategy for a specific circumstance will depend on several factors such as feedstock sources, conversion techniques, market opportunities, and especially the costs associated with these factors. Techno-Economic Analysis (TEA) provides an analytical framework to evaluate these tradeoffs and optimize biomass energy systems. Biomass feedstocks can be directly used as a solid fuels for heat and/or power production, or they can be transformed into liquid or gaseous fuels that are then suitable for a variety of applications.

Several biomass feedstocks can be used for bioenergy production including agricultural residues, municipal solid waste, food processing waste, animal waste, and dedicated energy crops. Dedicated energy crops can be divided into three categories: herbaceous energy crops that have little to no woody tissue, short-rotation woody crops, and oleaginous (lipid-rich) crops (Brown and Brown, 2014). Although, by definition, dedicated energy crops are not being grown for a food or feed application, some of these crops, such as corn (maize), sugar cane, sugar beet and soybeans, are also grown for food and livestock feed, which implies competition between food and energy markets. Growing both edible and non-edible energy crops often involves the conversion of food cropland and pasture to energy crops, and thus also affects food production through direct and indirect land use change, which collectively are increasingly referred to as induced land use change (ILUC). To avoid ILUC effects and a large carbon (C) debt, Robertson et al. (2017) suggests that perennial cellulosic crops should be planted on non-forested lands not already used for food production.

One benefit of cellulosic perennial energy crops is their ability to uptake nitrogen (N) at high rates throughout the growing season, and hence decrease leaching to groundwater (Robertson et al. 2017). Furthermore, the U.S. Congress has set a production target of renewable fuels by 2022 of 36 billion gallons (136 billion liters) of which 16 billion (60.5 billion liters) was intended to be in the form of cellulosic biofuel (EISA, 2007). Given the previously mentioned benefits, switchgrass, a perennial herbaceous energy crop~~,~~ shows potential as feedstock source for biomass energy systems. Although switchgrass has advantages over annual crops, the bioconversion process and logistics are challenging due to its high fiber content, low energy density, and varied properties.

The main macromolecular components of switchgrass and other lignocellulosic biomass feedstocks are lignin, cellulose, and hemicellulose. Among these, the most useful components for biological conversion to fuels are the cellulose and hemicellulose. However, access to these sugar polymers is hindered by several factors, including the presence of lignin. Lignin is a complex polyphenolic compound with many different molecular bonds. It is hydrophobic and forms a matrix around the cellulose and hemicellulose that is responsible for much of the rigidity of plant cell wall structure. These properties make lignin the hardest component of the plant cell wall to break down. To overcome the recalcitrance of lignocellulosic biomass, pretreatment is normally used to weaken the structure and thereby allow for access to the cellulose and hemicellulose.

Pretreatment can be accomplished by biological, chemical, thermochemical or mechanical processes. This step can be both energy and cost intensive. Thus, many research efforts have been dedicated to improving energy efficiency and reducing the costs related to the pretreatment process.

This study evaluates an alternative to pretreatment that represents a biomimetic approach to processing recalcitrant cellulosic biomass. This approach is modeled after a biological mechanism that has proven over time to efficiently deconstruct lignocellulosic biomass: the ruminant’s digestive system. In the case of anaerobic digestion (microbial conversion of biomass to methane), cotreatment may be a viable alternative to pretreatment. More specifically, mechanically disrupting the biomass while it is undergoing digestion, may have cost and process benefits. In many ways, this process mimics the design of a ruminant’s digestive system.

Chapter 2: Literature Review

2.1 Overview

Anaerobic digestion (AD) has been studied for many years. Recently, AD has been generating increased interest by academic and industrial researchers and innovators due to advances in technology, favorable policies, and improved economic feasibility. The following literature review summarizes relevant findings in the field of AD of lignocellulosic biomass, including government policies, feedstock sources, conversion techniques, market opportunities, and scale-up potential.

2.2 Anaerobic Digestion Science and Systems

Anaerobic digestion is a biological process in which microorganisms decompose organic feedstocks in the absence of oxygen. The main applications of AD include treating wastewater, municipal solid wastes, food wastes and agricultural wastes. The microorganisms involved in the decomposition process can be divided into several functional groups: hydrolytic, fermentative, acidogenic and methanogenic microorganisms (Adney et al., 1991). **Error! Reference source not found.** summarizes the main conversion steps taking place in anaerobic digestion (Adney et al., 1991; Rittmann and McCarty, 2001).

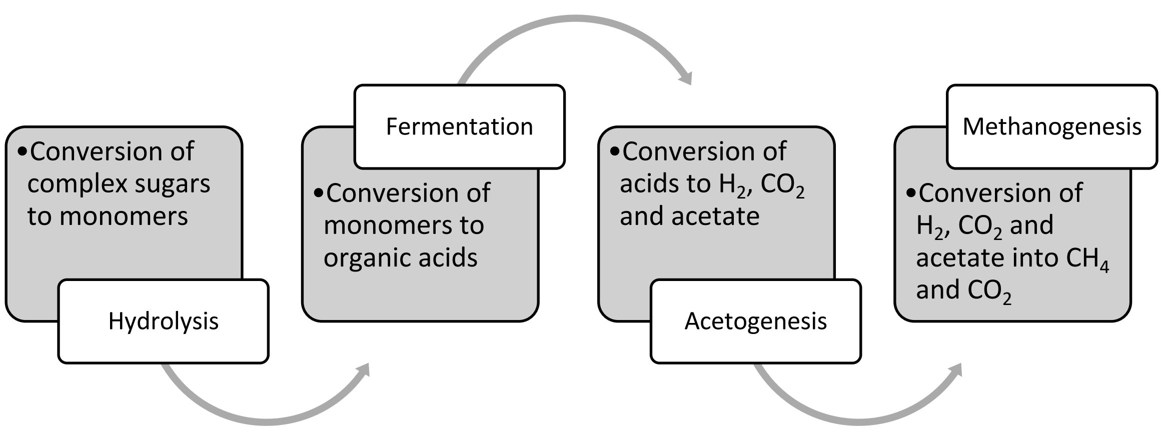


Figure 1: Main conversion steps of anaerobic digestion

First, the hydrolytic anaerobes convert complex organic matter like carbohydrates into simple monomers like sugars. Then the acidogens convert these monomers into organic acids. Acetogens convert these acids into acetate, carbon dioxide and hydrogen. Finally, methanogens use either acetate (acetoclastic) or hydrogen (hydrogenotropic) to produce biogas. Biogas is a gas mixture mainly composed of CO_2 and CH_4. Although they are presented in series, these steps can and generally do occur simultaneously in a reactor (Grady and Leslie, 2011). Stoichiometric reactions occurring in AD are summarized in **Error! Reference source not found.** and were used as the base for developing the stoichiometric equations modeled in the Aspen Plus model.

Table 1: Stoichiometric reactions of anaerobic digestion process

Reaction	Description	Extent of Reaction	Reference
$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$ $(C_5H_8O_4)_n + nH_2O \rightarrow nC_5H_{10}O_5$	Hydrolysis of complex sugars to monomers	0.4-0.7 of carbohydrate	(Liang et al., 2018; Rajendran et al., 2014)
$C_6H_{12}O_6 + 2H_2O \rightarrow 2CH_3COOH + 2CO_2 + 4H_2$	Acetogenesis	0.4±0.1 of sugar	(Rajendran et al., 2014)
$C_6H_{12}O_6 \rightarrow 3CO_2 + 3CH_4$	Methanogenesis	0.6 of sugar	(Rajendran et al., 2014)
$4H_2 + CO_2 \rightarrow CH_4 + H_2O$	Hydrogenotropic methanogenesis	1 of dihydrogen	Assumed since traces of H_2 or acetate were <0.1% found in experimental results (Liang et al., 2018)
$CH_3COOH \rightarrow CO_2 + CH_4$	Acetoclastic methanogenesis	1 of acetate	

Performance of AD will depend on several factors such as feedstock source, temperature, pH, reactor type and solids retention time. High solids content feedstocks (>20% total solids) are usually run on batch reactors, while feedstock containing 2-12% solids content can be run on continuously stirred tank reactors (CSTR) (Wellinger et al., 2013). Operating temperature ranges from mesophilic (30-35ºC) to thermophilic (50-60ºC) since at these temperatures the anaerobes thrive. Thermophilic conditions allow for a shorter retention times, the time a particle spends in a reactor (Chandra et al., 2012). Also the pH is maintained close to neutral (6.8-7.2) for the methanogens to survive. Some operating conditions for lignocellulosic biomass fed AD and performance are detailed in section 2.4.

Regulations and incentives have been influencing the steady growth of AD around the world. Europe has had great success in implementing AD at farm-scale scenarios for electricity generation and fertilizing due to a great price incentive (\$0.28 kWh^{-1}) provided for a period of 15 years(Dale et al., 2016). In the U.S. the renewable fuel standard has provided incentives as well to drive the production of renewable natural gas (Section 2.7). However, technological advances are needed to enhance the economics of AD plants (Vasco-Correa et al., 2018). Anaerobic biorefineries have been gaining interest for their ability to generate several products (heat, power, digestate) but TEA of these AD biorefinery concept are needed to evaluate the potential (Sawatdeenarunat et al., 2016).

2.3 Lignocellulosic Biomass as Feedstock for Anaerobic Digestion

2.3.1 Sources of Lignocellulosic Biomass

While its low digestibility limits its usefulness in traditional systems, lignocellulosic biomass still holds promise as a feedstock for biogas production. The DOE has estimated that crop residues and dedicated bioenergy crops are likely to be the primary sources of herbaceous lignocellulosic biomass in the future, while forest residues and short rotation woody crops are likely to be the primary sources of woody biomass. (Langholtz et al., 2016)

Lignocellulosic biomass can be produced by various dedicated bioenergy crops, which are crops not grown for food or feed applications. However, such crops can still impact food and feed production through induced land use change (ILUC). Literature suggests that planting perennial crops on economically marginal, non-forested lands that are not already used for food production can decrease ILUC effects and C debt (Robertson et al., 2017). Other benefits of perennial crops are related to water quality because they uptake Nitrogen (N) and other nutrients for a long growing season, and hence decrease nutrient leaching and runoff to water bodies (Somerville et al., 2010). Also, perennial crops require fewer inputs than annual crops and can be planted on marginal land (Werling et al., 2014). Due to these multiple benefits, perennial lignocellulosic biomass crops, such as switchgrass, may have considerable potential as feedstock sources for AD.

2.3.2 Lignocellulosic Biomass Structure

Lignocellulosic biomass is largely composed of three main macromolecular components: cellulose, hemicellulose, and lignin. The amount of each component varies depending on the biomass source. Some examples of composition of lignocellulosic biomass are reviewed in Sawatdeenarunat et al., 2015 (Table 2).

Table 2: Composition of several lignocellulosic biomass

Biomass	Cellulose (%)	Hemicellulose (%)	Lignin (%)	C/N Ratio	Reference
Corn stover	37.5	22.4	17.6	63	(Karthikeyan and Visvanathan, 2013; Li et al., 2014)
Wheat straw	38.2	21.2	23.4	60	(Brown et al., 2012; Karthikeyan and Visvanathan, 2013)
Switchgrass	31.0–45.0	20.0–31.0	12.0–18.0	90	(Brown et al., 2012; Karthikeyan and Visvanathan, 2013)
Bagasse	38.2	27.1	20.2	118	(Brown and Brown, 2014; Karthikeyan and Visvanathan, 2013)
Sugarcane	25	17	12	NA	(Karthikeyan and Visvanathan, 2013)
Rice straw	32	24	13	47	(Karthikeyan and Visvanathan, 2013; Ye et al., 2013)
Eucalyptus	38.0–45.0	12.0–13.0	25.0–37.0	NA	(Karthikeyan and Visvanathan, 2013)
Giant reed stalk	33.1	18.5	24.5	NA	(Monlau et al., 2012)
Giant reed leaves	20.9	17.7	25.4	NA	(Monlau et al., 2012)
Sunflower stalk	31	15.6	29.2	NA	(Monlau et al., 2012)
Biomass sorghum	22.2	19.4	21.4	NA	(Monlau et al., 2012)
Barley straw	37.5	25.3	26.1	NA	(Monlau et al., 2013)
Rye straw	38	36.9	17.6	20	(Monlau et al., 2013; Nizami et al., 2009)
Napier grass	45.7	33.7	20.6	26	(Janejadkarn and Chavalparit, 2013; Reddy et al., 2012)

Cellulose, the most common component, and hemicellulose are sugar polymers that can potentially be converted by microorganisms into methane, ethanol, and other fuels. The remaining component, lignin, presents the biggest challenge for biochemical processing and microbial conversion. To access the desired components, the recalcitrance of lignin must be overcome. Lignin is responsible for the rigidity and plant cell wall structure, and its complex polyphenolic bond structure makes it difficult to break down.

Once the lignin matrix is broken down, enzymes have much better access to cellulose and hemicellulose. To address this challenge of recalcitrance, a suite of biological, thermochemical, mechanical, and combinations of multiple pretreatment processes have been developed and investigated (Mosier et al., 2005; Tanjore and Richard, 2015).

Predictions of biogas yield require estimates of both theoretical yield, typically based on biomass composition and biofuel conversion stoichiometry, and actual yield as a function of biomass hydrolysis and conversion. Barbanti et al. (2014) reported theoretical biomethane yields of several lignocellulosic biomass sources (Table 3).

Table 3: Theoretical biogas yields of lignocellulosic biomass (Barbanti et al. 2014)

Biomass	CH_4 theoretical yield (mL g^{-1}VS)
Giant Reed	217 ± 5.3
Switchgrass	216 ± 5.5
Sorghum Silk	271 ± 5.9
Forage Hybrid Sorghum	251 ± 5.6
Fibre Sorghum	268 ± 5.5
Sweet Sorghum	256 ± 6.0
Maize	316 ± 3.7

2.4 Process Conditions for Anaerobic Digestion of Lignocellulosic Biomass

In considering how to implement cotreatment with anaerobic digestion, variables include reactor temperature and pH, solids loading rate, residence time, and the type of feedstock most likely to benefit. AD of high fiber, high solids content feedstocks can provide a higher organic loading rate, require a smaller reactor volume, decrease energy demand for heating, increase volumetric methane productivity, and generate less wastewater than conventional low solids AD (Yang et al., 2015). Table 4 summarizes some studies of lignocellulosic fed AD studies including process conditions and performance.

In addition to the previously described studies, Liang et al. (2018) reported the performance of mid-season switchgrass fed, thermophilic (55ºC) anaerobic digestion at different residence times.

Table 5 describes important properties of the process that are needed as model inputs and model validation. Although the study reports successful operation with residence times as low as 3.3 days, they did not include cotreatment.

Table 4: Review of lignocellulosic biomass fed AD studies

Feedstock	Operating Temperature	Solids Loading	Retention Time	Performance	Reference
Wheat straw	37ºC and 55ºC	2.5 $g_{VS}L^{-1}d^{-1}$	14-21 days	48% COD degradation, at thermophilic conditions, AD showed 36% higher methane yield than mesophilic	(Pohl et al., 2012)
Several lignocellulosic feedstocks	37ºC	18% (solid state) and 5% (liquid state) total solids	NA	No significant difference in methane yield between liquid and solid state, but the solid state showed 2 to 7-fold greater volumetric productivity than liquid state	(Brown et al., 2012)
Switchgrass	36ºC and 55ºC	20%-30% total solids	NA	30-53% cellulose removal, at thermophilic AD conditions, AD showed 30% higher methane yield than mesophilic	(Sheets et al., 2015)
Grass silage	35ºC	5.6%-11% initial solids	55 days	32-55% VS removal	(Lehtomäki et al., 2008)
Corn Stover	36ºC and 55ºC	20% total solids	NA	41% cellulose reduction, thermophilic conditions led to faster reductions of cellulose	(Shi et al., 2013)

Table 5: Properties of thermophilic anaerobic digestion of switchgrass for a 10 day residence time (Liang et al., 2018)

Fractional Carbohydrate Solubilization	0.644 ± 0.014
Biogas composition CO_2 : CH_4	47.5 ± 1.3 : 44.8 ± 1.1
Solids loading	30 g L^{-1}
Switchgrass composition:[1]	
Glucose	30.6% ± 1.4%
Xylose	22.9% ± 1.0%
Arabinose	4.07% ± 0.2%
Ash	5.92% ± 0.05%

2.5 Ruminant Digestive System Strategies Relevant to Anaerobic Digestion

Rumen digestion begins by grinding and moistening the ingested material before entering the rumen where it is inoculated with different microorganisms that both hydrolyze lignocellulose and ferment it into energy carrier molecules including carboxylic acids and methane. Repeated cycles of grinding ("chewing the cud") and additional digestion are estimated to occur hundreds of times for each mouthful of biomass. Weimer's (2009) review highlights the efficient use of energy of cattle to digest lignocellulosic biomass (<2% of the energy consumed in their feed). To date no commercialized bioconversion system has come close to achieving the digestive performance of ruminants in terms of conversion rates, yields, or energy efficiency. Research suggests that in addition to the environmental conditions (pH, temperature, microbial community, etc.) of the rumen, the mechanical strategies of ruminant fermenters should be mimicked to improve anaerobic digesters (Bayané and Guiot, 2011).

Weimer et al. (2009) proposed an integrated process that would consist of an anaerobic biodigester with continuous feedstock flow and milling between cycles of digestion (Figure 2). The rumen is where biomass breaks down biologically and chemically through fermentation and microbial activity. This process is analogous to anaerobic digestion. The mouth is where biomass breakdowns mechanically through chewing. This process is analogous to milling. Finally, the esophagus connects the mouth with the rumen for biomass transportation. This process is analogous to pumping. This cycle occurs several times until biomass is mostly digested and follows through the rest of the digestive process. The concept of combining fermentation with milling of partially digested biomass has been termed

[1] Feedstock composition also reported in DiMarco, 2017

"cotreatment" (Paye et al., 2016). The implementation of such mechanical cotreatment with digestion, simulating the features of grass-fed ruminants like cattle, may be able to decrease both the energy and the cost of the process (Weimer et al., 2009)

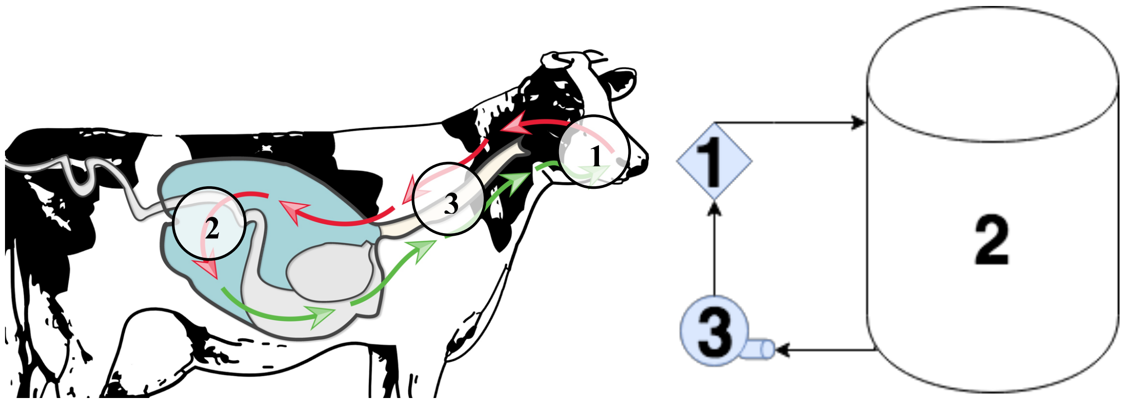


Figure 2: Ruminant digestive system steps (chewing, transportation to rumen and digestion) and analogous steps in an engineered systems (mechanical disruption, transportation to digester and AD)

2.6 Cotreatment of lignocellulosic biomass as a strategy

2.6.1 Cotreatment Integration with Pure Culture Fermentations

Consolidated bioprocessing (CBP) combines enzyme production, enzymatic hydrolysis, and fermentation in a single reactor. It has many economic advantages since it reduces the cost of biological conversion significantly in comparison to other processes (Lynd et al., 2005). Some of the factors responsible for the cost reductions are avoided capital costs and capital investment, utilities, and reduced reactor volume. Cotreatment and strategies of mechanical disruption during fermentation have previously been suggested as a way to further improve CBP in pure culture fermentations (Weimer et al., 2009; Bayané and Guiot, 2011), but until recently there was no experimental demonstration of its performance.

Recent results with pure-culture fermentations of *Clostridium thermocellum*, a CBP organism, show significant improvements in biomass solubilization, gas yield, and particle size reduction for cotreatment integrated to CBP (Balch et al., 2017; Paye et al., 2016). Anaerobic pure culture fermentations of lignocellulosic biomass with integrated cotreatment have also shown improvement in biogas yield (Lindner et al., 2015), particle size reduction (Paye et al., 2016) and biomass solubilization due to increased accessibility of plant cell walls to biological attack (Balch et al., 2017; Paye et al., 2016). Figure 3 shows an increase of more than 40% carbohydrate solubilization with cotreatment when compared with no cotreatment. Balch et al. (2017) also reports no detrimental effect of milling on *Clostridium thermocellum,* in comparison, it prevented entirely the soluble substrate fermentation by yeast.

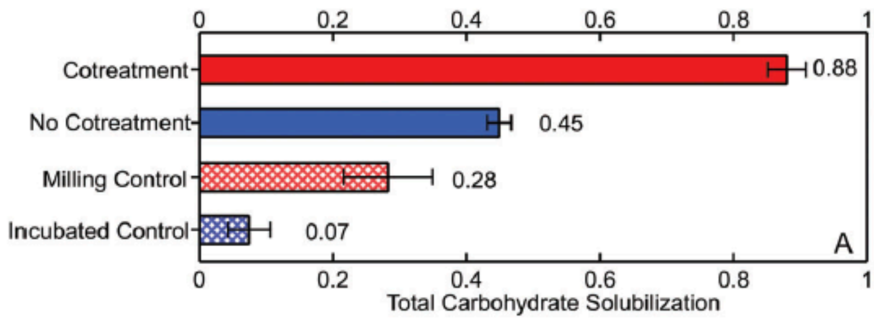


Figure 3: Carbohydrate solubilization by *Clostridium thermocellum* of mid-season switchgrass. Treatments included CPB with cotreatment, CBP with no cotreatment, milling prior to fermentation without inoculation with *C. thermocellum*, and no inoculation or pretreatment (Balch et al., 2017)

Adding cotreatment to CBP and other technology advances is projected to be a promising strategy that can provide lower costs when compared to other existing processes (Lynd et al., 2017). Figure 4 shows an eightfold shorter payback period reduction when a full suite of advanced technologies are implemented (cotreatment, CBP and process optimization) in comparison to a base case (dilute and acid pretreatment and fungal cellulase addition previous to fermentation).

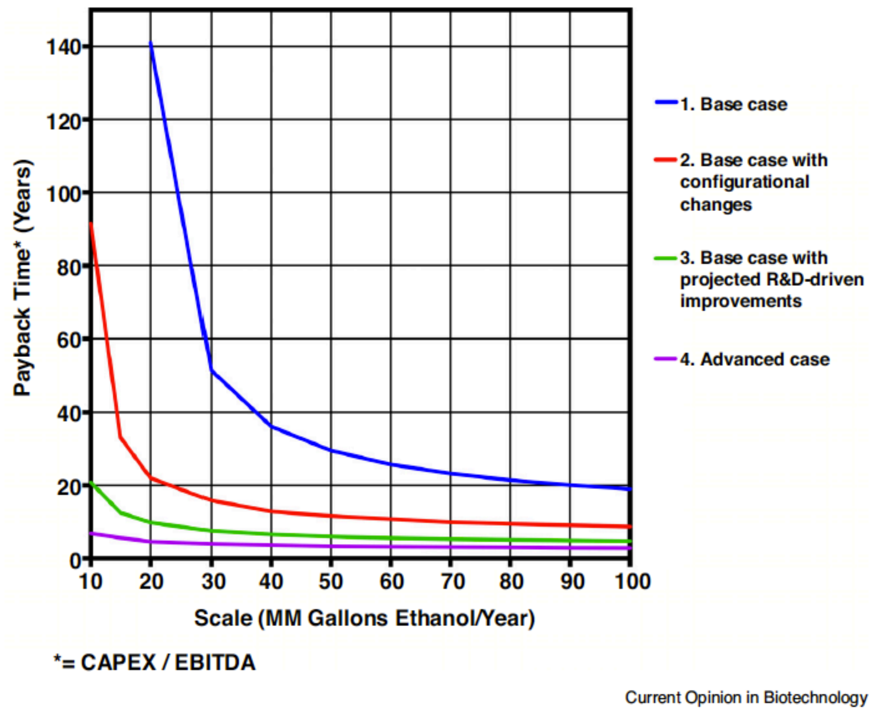


Figure 4: Payback time of base to advanced technology scenarios at different scales (Lynd et al., 2017)

Many of the previous studies converge in their recommendations for future research needs. These include a need to characterize energy requirements for milling in cotreatment, techno-economic analysis of processes to assess industrial scale feasibility, and the development of efficient handling of high fiber solid content feedstocks.

2.6.2 Cotreatment Integration with Mixed Culture Fermentations

Anaerobic digestion with mixed cultures is a traditional form of CBP since enzyme production, hydrolysis, fermentation and methanogenesis occur in the same reactor. Some studies have investigated the effect of mechanical pretreatment on AD as well as two-stage digestion with intermediate treatment between stages and are summarized in Table 6.

In general, mechanical treatment in between digestion has shown significant improvements in methane yield and production rates. Initial results suggest implementing cotreatment to anaerobic digestion should be quite promising (Bharadwaj, 2019). However, further research related to these initial findings, including techno-economic modeling, is required before investors and engineers will have the confidence to develop full-scale bioenergy processes.

Table 6: Summary of mechanical and intermediate treatment studies in AD

Feedstock	Treatment Description	Effect on AD	Reference
Municipal Solid Waste	Pressurized steam treatment (240ºC) in between two-stage AD	Doubled CH_4 production rate and increased CH_4 yield by 40%	(Liu et al., 2002)
Maize Silage	Ball milled and placed into sample bottles	9% methane production increase	(Lindner et al., 2015)
Hay/straw digestate	Ball milled and placed into sample bottles	17% methane production increase	(Lindner et al., 2015)
Digestate from full scale digester	Ball milled and placed into sample bottles	Quadruplication of daily methane production	(Lindner et al., 2015)
Sewage Sludge	Intertreatment: thermal steam explosion between two-stage AD	Improvements in methane production of 20%-40%	(Ortega-Martinez et al., 2016)

2.7 Relevance of U.S. Bioenergy Policies

The US government has established several laws and policies related to the production of renewable energy (EISA, 2007; U.S. Congress, 2005). These policies have been driven by 3 main factors: national security, energy supply, and environmental sustainability (Bracmort, 2015; Brown and Brown, 2014; EISA, 2007; U.S. Congress, 2005). In the past, national energy security and supply have been viewed as critical concerns, and threatened by international events and predicted shortages of oil and gas supplies. But in the last few years these issues have receded from public attention due to technology advancements in hydraulic fracturing, which provides the U.S. with an expanding supply of low-cost natural gas and shale oil (Kilian, 2016). However, environmental sustainability remains a global concern because of climate change impacts attributed to carbon emissions (Pachauri et al., 2014). Biobased energy sources are relevant because they have the potential to decrease carbon dioxide (CO_2) and other greenhouse gas (GHG) emissions in comparison to fossil fuel sources. Biobased energy, if done right, can also enhance nutrient cycling and soil and water quality in agricultural systems (Dale et al., 2016).

The most recent federal government mandate related to biobased energy production in the United States is the Renewable Fuel Standard (RFS), amended as part of the Energy

Independence and Security Act (EISA, 2007). EISA established a target production rate of 36 billion gallons (136 billion liters) of renewable fuels by 2022. More specifically, Congress set a target that 16 billion gallons (60.5 billion liters) out of that overall goal should be in the form of cellulosic biofuel. Congress authorized the U.S. Environmental Protection Agency (EPA) to administer the effort through a market-based system that requires fuel blenders to buy renewable fuel credits (Bracmort, 2015). These markets for Renewable Identification Numbers (RINs) are intended to encourage the production of renewable fuels, and include different RIN classifications based on feedstocks and the relative reductions of GHG emissions relative to gasoline. In average, RIN selling price in 2016 was \$2.24 gallon of ethanol^{-1} (Environmental Protection Agency, 2018).

In addition to these RIN subsidies, one of the critical factors that determines commercial investment is the predicted price of production of different renewable fuel types from different feedstocks. Gasoline gallon equivalent (GGE) is a unit often used for comparison of alternative fuels with fossil fuels (U.S. Department of Energy, 2014a). For example, using pilot-scale validation data, the U.S. Department of Energy modeled a technology pathway for hydrocarbon biofuel production at a price of \$3 GGE^{-1} (2014 USD) to demonstrate a viable commercialization pathway (U.S. Department of Energy, 2016a). In contrast, a gasoline gallon was priced at \$3.43 and natural gas \$2.17 GGE^{-1} in 2014 (U.S. Energy Information Administration, 2019). However, risks associated with new technologies, the large capital investments required to achieve economies of scale, and the uncertainties of future subsidies (RINs) and baseline fossil fuel prices all have limited commercial production of cellulosic biofuels to levels greatly below the RFS2 Congressional goals (Lynd, 2017).

Although the RFS targets originally assumed that biofuels would be in liquid forms such as ethanol or drop-in fuels, the growing use of fossil natural gas for transportation has opened an opportunity for Renewable Natural Gas (RNG) to also count as a cellulosic biofuel. In 2014 the EPA authorized the use of RNG to count toward RFS goals, and since then RNG has been responsible for over 95% of cellulosic biofuel production in the U.S (Figure 5).

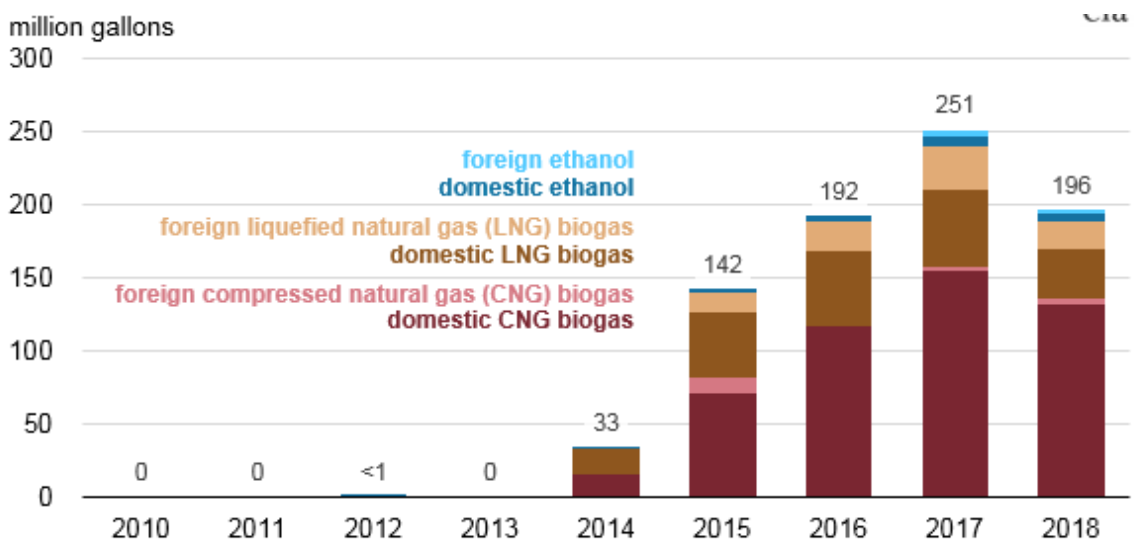


Figure 5: Renewable fuel standard cellulosic biofuel production from 2010-2018 (Source: US Energy Information Administration based on RFS Program)

To date this RNG has been almost entirely from very large sources, initially biogas from solid waste in landfills and more recently from manure at livestock facilities. The economies of scale associated with conventional AD systems and for biogas upgrading to pipeline-quality RNG have thus far limited small and mid-sized farms from benefiting from these incentives (Leuer et al., 2008). Adding dedicated energy crops to manure can potentially allow larger AD operations on farms with smaller numbers of livestock.

2.8 Techno-economic Analysis

Techno-economic analysis (TEA) is a tool used to assess the technical and economic performance of a plant through a specific pathway (Brown and Brown, 2013). It is an inexpensive way of assessing the potential of naïve technologies on a commercial scenario. It has become very useful to evaluate biorenewable pathways that can be successful and competitive with existing markets.

The National Renewable Energy Laboratory (NREL) has published many TEAs related to pure culture biorefineries (Davis et al., 2014; Humbird et al., 2011; Zhu et al., 2014). These include useful information on biomass physical properties (Wooley and Putsche, 1996) for process modeling as well as capital and operating costs for plant economic analysis. There is a TEA simulation of cotreatment in pure culture fermentations (Lynd et al., 2017), but not of mixed culture anaerobic microbiomes.

An initial TEA of pure culture fermentations that include cotreatment integrated with CBP indicates that this approach can significantly reduce costs for biological conversion (Lynd et al. 2017a). However, even in the pure-culture case there are still questions about industrial scale-up and economic viability. Most current cellulosic conversion technologies have a payback period on just the capital investment of about two decades, but Lynd et al. (2017) estimates that cotreatment has a much more promising short-term payback period of less than a decade even at relatively small scale. However, the Lynd et al. (2017) study states the need for accurate data regarding milling energy requirements to document both conversion efficiency and operating costs. Techno-economic analysis of cotreatment integrated with CBP can help answer this and other scale-up viability questions. These questions can also be informed by previous studies that perform techno-economic analysis of bioenergy processes, and can provide a framework for comparing the results of cotreatment aided AD with other bioenergy technologies (Zhu et al., 2014).

Chapter 3: **Goals, Objectives and Hypotheses**

The following sections describe the goals, objectives and hypotheses of the proposed research study that will contribute to understanding and addressing the challenges previously stated.

3.1 **Goal**

The proposed research intends to use techno-economic analysis to assess the impact of cotreatment on lignocellulosic biomass-fed anaerobic digestion under thermophilic conditions. The intent of the research is to compare both scenarios in terms of mass flows and costs.

3.2 **Objectives**

Assuming cotreatment has a similar improvement in fractional carbohydrate solubilization for mixed culture fermentations, develop TEA assessment of two commercial scenarios of switchgrass fed, thermophilic, AD commercial system process with and without cotreatment in order to:

A) Compare mass (kg hr^{-1}) and energy flows (MMkcal hr^{-1}) of both scenarios through process modeling.

B) Compare capital expenses (USD 2014), operating expenses (USD 2014) and minimum fuel selling price ($ GGE^{-1}) of both scenarios through a discounted cash flow analysis.

C) Determine the effect of fractional carbohydrate solubilization (%) and scale in process costs ($ GGE^{-1}) through sensitivity analysis.

3.3 **Research Questions**

A) What potential percentage increase in product flow would cotreatment allow relative to a no cotreatment scenario?

B) How does cotreatment affect the costs of anaerobic digestion of lignocellulosic biomass under thermophilic conditions?

C) How do carbohydrate solubilization and scale affect the costs of lignocellulosic biomass-fed anaerobic digestion under thermophilic conditions?

D) What are the minimum scale and solubilization parameters under which thermophilic anaerobic digestion of lignocellulose is economically feasible in comparison to natural gas?

Chapter 4: Methodology

4.1 Overview

The intent of the study is to and assess the scale-up potential of cotreatment for lignocellulosic biomass fed anaerobic digestion through techno-economic analysis. Thus, the study was divided in two phases: process modeling and economic analysis. The results from the process modeling phase served as input to the economic analysis phase. Both phases

Figure 6) are described further in the following sections. Two scenarios were modeled: without cotreatment and with cotreatment.

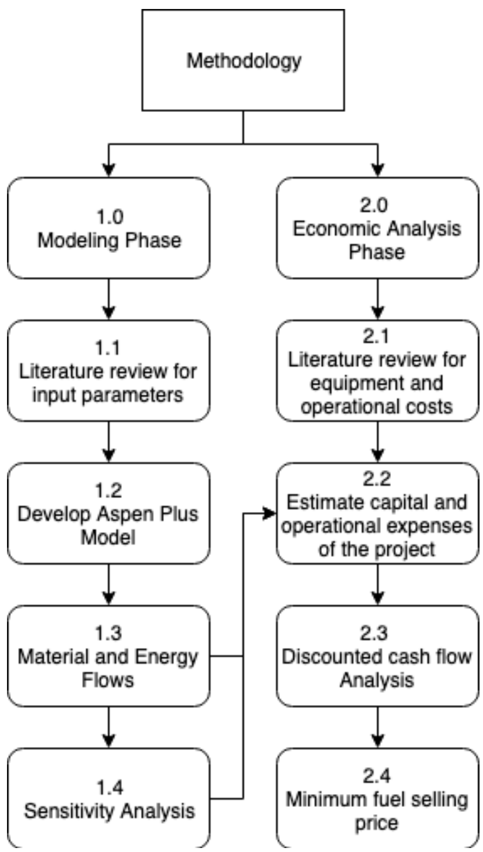


Figure 6: Methodology approach

4.2 Modeling Phase

4.2.1 Description

The modeling phase consisted of developing a process model for both scenarios, one with cotreatment and the other without cotreatment. The goal of this phase is to obtain material and energy flow data for each section and scenario that will serve as inputs for the economic analysis.

4.2.2 Equipment & Materials

The process simulations were modeled using Aspen Plus® V10 (36.0.0.249) software. The software is available in the Agricultural and Biological Engineering Department computer lab. mass and energy flows were obtained from the simulation results, calculated by stoichiometry and thermodynamic laws. The mass flows will be reported in kg hr^{-1} units and the energy flows are reported in the form of enthalpy flows of each stream in MMkcal hr^{-1}.

Process economics were estimated afterward by incorporating the results of the Aspen Plus models into an Excel spreadsheet. Data from published literature provided most necessary process inputs, supplemented by unpublished experimental data from the Richard lab when available. The physical properties were adapted from Humbird et al. (2011) and are described in Appendix A: Aspen Plus Properties.

4.2.3 Model Inputs and Assumptions for Each Scenario

Table 7: Summary of process model input assumptions

Fractional Carbohydrate Solubilization	No cotreatment: 0.644 ±0.014
	Cotreatment: 0.819± 0.125
Feedstock Composition	Mid-season switchgrass
Glucan	36.44 %
Xylan	24.54 %
Arabinan	2.91 %
Galactan	1.12 %
Lignin	2.92 %
Feedstock input flow	2,000 dry Mg day^{-1}
Solids Content (input)	6% total solids
Digester pH	7
Digester Retention Time	10 days
Scale Sensitivity Analysis (dry Mg day^{-1}**)**	20, 110, 200, 1100, 2000
Cotreatment Energy Consumption	72 MJ Mg^{-1} (dry basis)

4.2.4 Model Description

Figure 7 illustrates a schematic of the process overview to be modeled. The process is divided into 4 areas (A100, A200, A300, and A400) , all of them described in detail below.

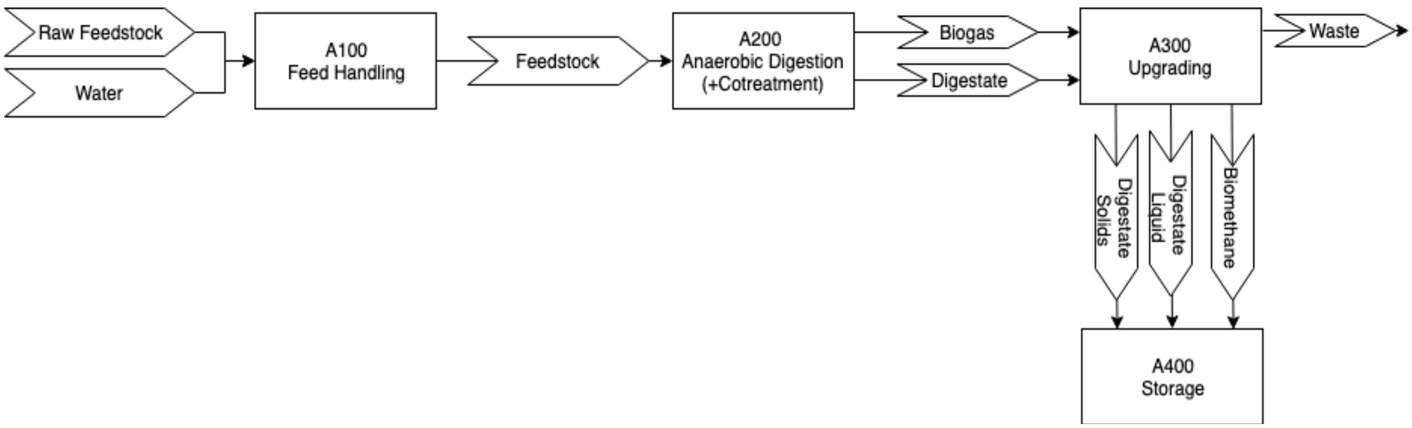


Figure 7: Process overview of the anaerobic digestion commercial scenario modeled in Aspen Plus

Area 100: Feed Handling

This area consists of feed reception, in this case mid-season switchgrass (Table 8: Quantitative saccharification of switchgrass, mixing with water, and conveyance to the next area. The feedstock input is 2000 dry Mg day^{-1} and the total solids content is 6% (60gL^{-1}). Figure 8 shows the flowsheet for the feed handling module (Area 100). Stream 101 is water, stream 102 is switchgrass and stream 103 is mixed water and switchgrass. A mixing block is used with an outlet pressure of 1 atm to combine the input streams.

For this study, the composition is based on analysis of mid-season switchgrass, as previously analyzed by the Richard lab (DiMarco, 2017) and presented in Table 8.

Table 8: Quantitative saccharification of switchgrass used in this study

Sample	%Sucrose	%Water Extractable Others	%Ethanol Extractives	% Lignin	% Glucan	% Xylan	% Galactan	% Arabinan	Total %
ERNST Switchgrass	0.09	4.96	2.52	22.92	36.44	24.54	1.12	2.91	99.39

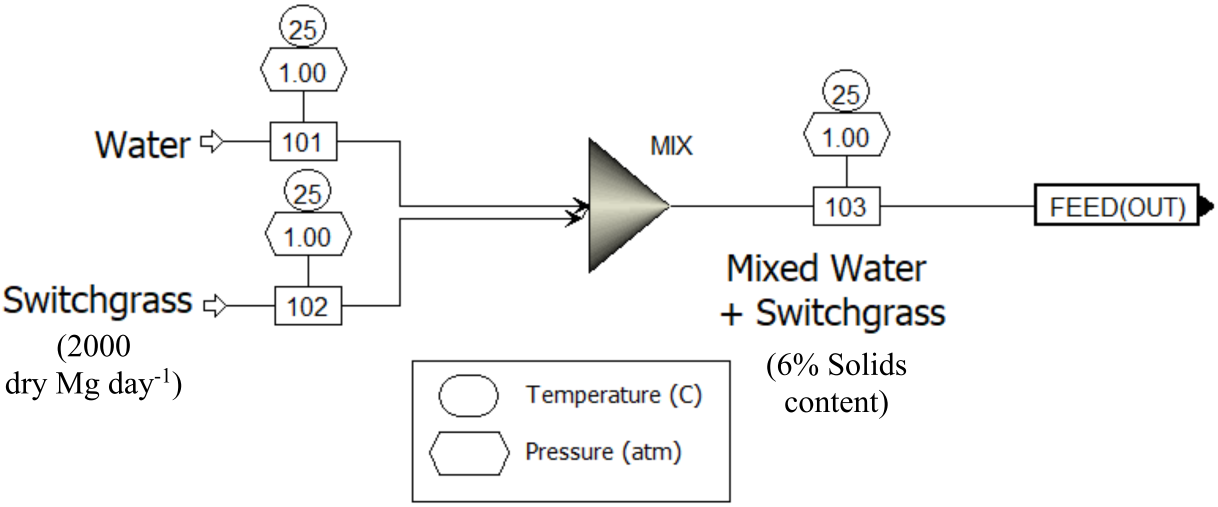


Figure 8: Area 100 flowsheet for feed handling

Area 200: Anaerobic Digestion (with Cotreatment)

This area consists of the anaerobic digestion process and a cotreatment recirculation loop in the case of the cotreatment scenario. The gas and liquid flows are then divided into separate streams. The digester was modeled at thermophilic temperature (55ºC) for a residence time of 10 days. Figure 9 shows the process flowsheet for area 200. Stream 103 is the feedstock input, stream 201 is the feedstock being pumped to the reactor, stream 202 is the output stream containing both gas, liquid, and solid phases. The flash block then separates the biogas (stream 204) from the solids and liquids (digestate).

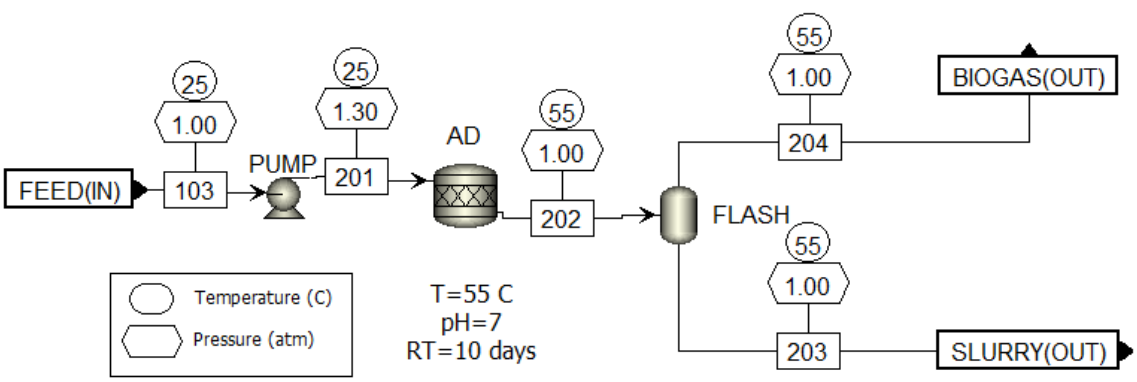


Figure 9: Area 200 flowsheet for anaerobic digestion

Table 9: Stoichiometric Reactions of Reactor Block in Area 200

Reaction Number	Fractional Conversion	Fractional Conversion of Component	Stoichiometry	Description
1	0.644± 0.014	Cellulose	$(C_6\ H_{10}\ O_5)n\ +\ nH_2O \rightarrow nC_6\ H_{12}\ O_6$	Hydrolysis of cellulose to glucose
1c	0.819± 0.125			
2	0.400	Glucose	$C_6H_{12}O_6 + 2H_2O \rightarrow 2CH_3COOH + 2CO_2 + 4H_2$	Acetogenesis
3	0.600	Glucose	$C_6H_{12}O_6 \rightarrow 3CO_2 + 3CH_4$	Methanogenesis
4	0.644± 0.014	Xylan	$(C_5\ H_8\ O_4)n\ +\ nH_2O \rightarrow nC_5\ H_{10}\ O_5$	Hydrolysis of xylan to xylose
4c	0.819± 0.125			
5	0.400	Xylose	$C_5H_{10}O_5 + H_2O \rightarrow 2CH_3COOH + CO_2 + 2H_2$	Acetogenesis
6	0.600	Xylose	$2C_5H_{10}O_5 \rightarrow 5CO_2 + 5CH_4$	Methanogenesis
7	0.644± 0.014	Arabinan	$(C_5\ H_8\ O_4)n\ +\ nH_2O \rightarrow nC_5\ H_{10}\ O_5$	Hydrolysis of arabinan to arabinose
7c	0.819± 0.125			
8	0.400	Arabinose	$C_5H_{10}O_5 + H_2O \rightarrow 2CH_3COOH + CO_2 + 2H_2$	Acetogenesis
9	0.600	Arabinose	$2C_5H_{10}O_5 \rightarrow 5CO_2 + 5CH_4$	Methanogenesis
10	0.644± 0.014	Galactan	$(C_6\ H_{10}\ O_5)n\ +\ nH_2O \rightarrow nC_6\ H_{12}\ O_6$	Hydrolysis of galactan to galactose
10c	0.819± 0.125			
11	0.400	Galactose	$C_6H_{12}O_6 + 2H_2O \rightarrow 2CH_3COOH + 2CO_2 + 4H_2$	Acetogenesis
12	0.600	Galactose	$C_6H_{12}O_6 \rightarrow 3CO_2 + 3CH_4$	Hydrogenotropic methanogenesis
13	1	Dihydrogen	$4H_2 + CO_2 \rightarrow CH_4 + H_2O$	Acetoclastic methanogenesis
14	1	Acetic Acid	$CH_3COOH \rightarrow CO_2 + CH_4$	Methanogenesis

Table 9 describes the stoichiometry modeled for the reactor. Since there are no published data for cotreatment effect on mixed culture anaerobic digestion, the fractional conversion of structural carbohydrates for the cotreatment scenario was varied from an increase of 5% to 30% in comparison to the no cotreatment scenario. No recirculation loop is modeled in Aspen Plus, but the equipment and energy consumption was taken into account in the economic analysis. The reaction numbers followed by a “c” indicate conversion used for cotreatment scenario.

<u>Area 300: Upgrading</u>

This area consists of separating the gas substreams to upgrade biogas into pipeline quality 98% biomethane. Figure 10 shows the overview of Area 300. First, the water (stream 307) is removed from the biogas in the 300SEP block by setting the H_2O split fraction to 1 on stream 307 and the split fraction of the biogas to 1 on stream 308. Afterwards, the biogas (stream 308) is compressed (300 Comp) to 13atm and enters the membrane separation system (modeled with 301SEP block). Here, CO_2 and other gases are removed from the biogas by setting the split fraction of these gases to 1 and 0.2 for the CH_4 stream 304. The split fraction for CH_4 (Stream 303) is set to 0.98. In this study the remaining gases are considered waste streams (stream 304), although the CO_2 stream could potentially have value for utilization or geologic sequestration and significantly improve the GHG balance of the process. Lastly, the 98% CH_4 stream is compressed to 250 atm (block 301COMP) and cooled down to 25ºC, which are the conditions at which compressed natural gas (CNG) is usually stored at. The membrane lifespan is 6 years (Bauer et al., 2013). All compressors are assumed isentropic with a 72% efficiency.

In this area the digestate is pumped , separated into solids and liquids and transferred to storage (Fuchs and Drosg, 2013).

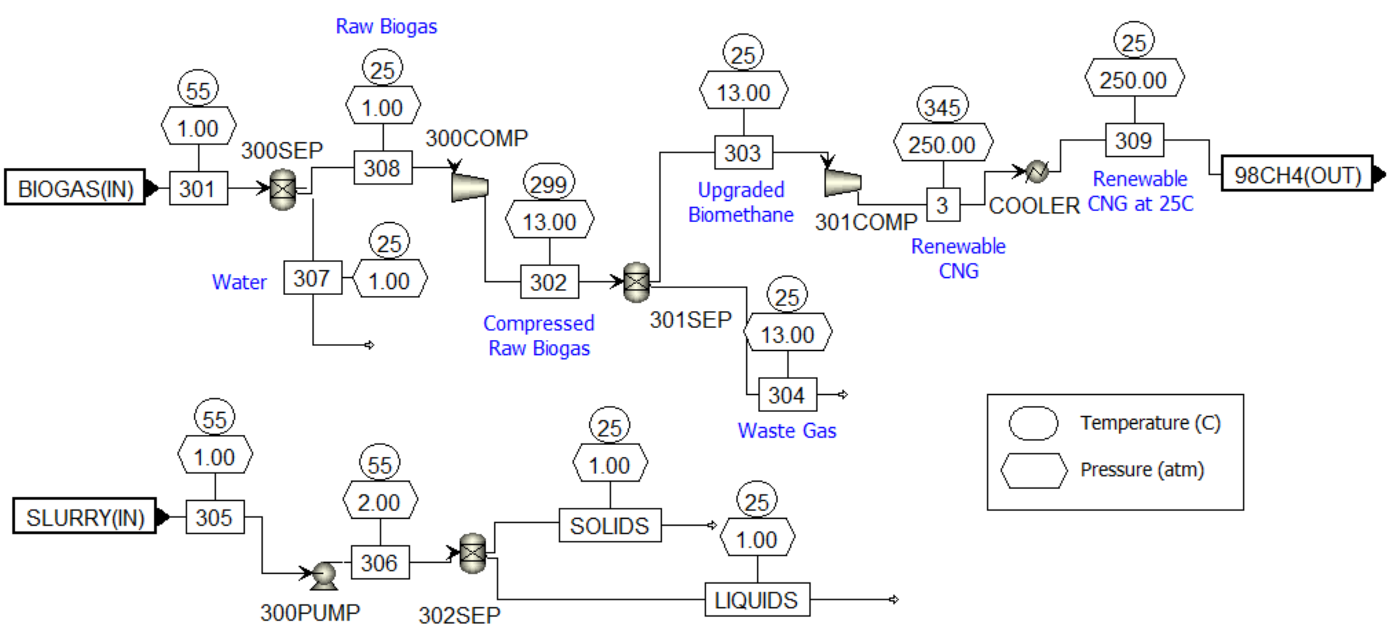


Figure 10: Area 300 flowsheet for upgrading

Area 400: Storage

This area collects all the digestate and biomethane and stores it for later sale. This aspect of the system logistics is not modeled in Aspen Plus since there are no physical changes to the product properties. However, the equipment and operating costs are estimated for the economic analysis (Smith and Gonzales, 2014).

4.2.5 Sensitivity Analysis

The results of the model include mass flows and energy flows for each stream as well as energy consumption of each block. A sensitivity block was modeled for the uncertainty in the fractional conversion of carbohydrates. This sensitivity block is a model analysis tool in Aspen Plus that allows the user to vary model parameters according to a list of values, and equidistant distribution or a logarithmic one. This variable was chosen since results for cotreatment of mixed culture systems are not yet available, and is based on assumptions that cotreatment will have similar effects as it has had in pure culture systems. The 98% CH_4 product mass flows was then compared between scenarios to explore the effect of carbohydrate solubilization on biomethane production. Equation 1 was used to determine the energy conversion efficiency of each scenario. The high heating value (HHV) of switchgrass was assumed to be 18 MJ kg^{-1} (Ragauskas et al., 2010) and the HHV of CNG was assumed 47 MJ kg^{-1} (U.S. Department of Energy, 2014b)

$$\frac{E_{biomass}}{E_{product}}(\%) = \frac{Energy\ in\ RNG\ stream\ (\frac{MJ}{hr})}{Energy\ in\ feed\ \left(\frac{MJ}{hr}\right)} \qquad \text{(Eq. 1)}$$

Another sensitivity analysis was performed for the scale of the process. This sensitivity analysis considered 5 different scales (dry Mg day^{-1}): 20, 110, 200, 1100, and 2000. The purpose is to evaluate at which scales the process is economically feasible.

4.2.6 Model Validation

The model was validated using Liang et al. (2018) published data as input. The experimental results reported a steady state value for biogas production rate (mL day^{-1}) and CO_2 : CH_4 ratio. The model was run with the same inputs as the experiment: 100mL day^{-1} feed for a 10 day retention time, at 30g L^{-1} (3%) solids content and 0.644±0.014

fractional carbohydrate solubilization. The results of the model run were compared to the study's reported results and a percentage difference was calculated using Eq. 2.

$$\%Diff = \frac{R_{modeled} - R_{experimental}}{(R_{experimental})} * 100 \qquad \text{Eq. (2)}$$

4.3 Economic Analysis Phase

4.3.1 Description

The aim of economic analysis phase goal is to determine a minimum fuel selling price based on the material and energy flows from the model for both scenarios. The minimum fuel selling price is the price ($ GGE^{-1}) at which biomethane can be sold to obtain a net present value of zero, after a 10% internal rate of return after taxes as the discount rate.

4.3.2 Equipment & Materials

An Excel spreadsheet based on NREL's previously published TEA models was used for the economic analysis (Humbird et al., 2011). The spreadsheet imports the process model material flows through Visual Basic (VBA) coding.

4.3.3 Assumptions

The metric for cost estimation was the minimum fuel selling price (MFSP), the field (plant) to fuel (output gate) price that will generate a net present value (NPV) of 0 over the lifetime of the modeled plant. This metric is chosen as it provides a useful comparison to other TEA studies of renewable fuels as well as market prices for fossil fuels, and can also be used to assess cost-competitiveness of renewable natural gas. An n^{th} plant assumption is made, reflecting a mature future with a successful industry of many established plants.

Discounted cash flow analysis was used to find the MFSP at which NPV=0. Economic analysis results are reported in 2014 USD to be in comparable with DOE goals (U.S. Department of Energy, 2016). Feedstock cost was assumed to be $77.60 dry Mg^{-1} based on these same DOE targets. The parameters for the discounted cash flow analysis were adapted from Humbird et al. (2011) and are summarized in Table 10.

Table 10: Discounted cash flow analysis parameters

Plant life	30 years
Discount rate (Internal Rate of Return)	10%
General plant depreciation	200% declining balance (DB)
General plant depreciation recovery period	7 years
Steam Plant Depreciation	150% DB
Steam Plant depreciation recovery period	20 years
Federal tax rate	35%
Financing	40% equity
Loan terms	10-year loan at 8% APR
Construction period	3 years
First 12 months expenditures	8%
Next 12 months expenditures	60%
Last 12 months expenditures	32%
Working capital	5% of fixed capital investment
Start-up time	3 months
Revenues during start-up	50%
Variable costs incurred during start-up	75%
Fixed costs during start-up	100%
Feedstock Cost	\$77.60 Mg^{-1} (dry basis)

The site development, warehouse and piping costs are based on the inside-battery-limits equipment costs (Humbird et al., 2011). The inside-battery-limits (a systems boundary term used in the refinery and manufacturing sectors) are considered part of the total direct costs of the plant and includes all costs associated with procuring and installing all process equipment.

4.3.4 Discounted Cash Flow Analysis and Sensitivity

Capital and operating costs were estimated using prices and quotes previously reported in literature (Humbird et al., 2011). Plant cost indices (Chemical Engineering, 2019), labor index (U.S. Department of Labor and Bureau of Labor Statistics, 2017) and electricity cost were used for proper year-dollar back-casting using Eq. 3 (Humbird et al., 2011). The detailed equipment, operating costs and indices are detailed in Appendix B: Equipment Costs, Appendix C: Operating Costs, Appendix F: Cost Indices respectively.

$$2014\ Cost = BaseCost * \frac{2014\ Cost\ Index}{Base\ Year\ Index} \quad \text{(Eq. 3)}$$

The equipment was sized accordingly for the desired size based on mass flow streams by using Eq. 4 (Brown and Brown, 2014) and are described further in Appendix B: Equipment Costs.

$$C_{p,s} = C_{p,s} \left(\frac{S_s}{S_b}\right)^n \qquad \text{(Eq. 4)}$$

where,

$C_{p,s}$ =predicted cost of equipment

$C_{p,s}$ =known cost of base equipment

S_s =desired size of equipment

S_s =known size of base equipment

n =economy of scale sizing exponent

The sizing exponents are based on previous literature and are less than 1. The sizing exponents used for each equipment are detailed in (Appendix B: Equipment Costs). The only sizing exponent not available in literature is for the colloid mill being tested for cotreatment. Personal communication with an IKA company sales representative was established to gather mill quotes at pilot and commercial scales (Broadwater, 2018). From these quotes, a scaling factor based on power consumption was calculated (Appendix E: Mill Scale-up Calculations).

Preliminary milling tests have been performed that estimate milling energy to be 72 MJ dry Mg^{-1} herbaceous biomass (Bharadwaj, 2019). This value was used for the cotreatment part of the economic analysis. Based on the feed input flow, the energy spent on cotreatment per day was calculated. For the biogas upgrading equipment costs were used from Bauer et al. (2013) which contained information on both capital and operating expenses.

For the discounted cash flow analysis, all annual discounted cash flows were summed to calculate the NPV. The annual discounted cash flow was calculated with Eq. 5:

$$A_{DCF} = \frac{A_{CF}}{(1+i)^n} \qquad \text{(Eq. 5)}$$

where,

A_{CF} =annual revenues less operating expenses, capital investment and federal income tax

n =year

i =discount rate

The ultimate result will be estimated as MFSP ($ GGE^{-1}). A sensitivity analysis was performed to evaluate carbohydrate solubilization impact on MFSP as well as scale impact on MFSP.

4.3.5 Validation of Economic Model

The MFSP calculated will be compared to existing TEA results of renewable natural gas projects and/or models. The MFSP will also be compared to current natural gas prices and RIN incentives.

Chapter 5: Results & Discussion

5.1 Process Modeling Phase

The following section presents mass flows results of the process model for the two process scenarios broken down by process area.

5.1.1 Process Model Validation Results

Table 11 shows the model validation results. The model output has a 6.1% higher biogas production rate and a 6.1% higher CH_4 molar ratio. The percent difference is less than 10%.

Table 11: Model validation results

	Biogas production rate (mL day^{-1})	**CO_2 (molar ratio)**	**CH_4 (molar ratio)**
Model Results	957.3 ± 21.2	49.3	47.6
Experimental	928.8 ± 51.1	47.5	44.8
%Diff	3.1%	3.8%	6.3%

5.1.2 Area 100: Mass and energy flows for feed handling

Table 12 shows results for the feed handling area of the model for both scenario A (no cotreatment) and scenario B (cotreatment). Table 13 shows results for the energy flows. This result was expected since the same inputs were received for both simulated AD plants, and were preprocessed the same way.

Table 12: Area 100 mass flows of feed handling

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Raw Feedstock	90,000	kg hr^{-1}	Raw Feedstock	90,000	kg hr^{-1}
	Water	1,291,667	kg hr^{-1}	Water	1,291,667	kg hr^{-1}
OUT	Feedstock	1,381,667	kg hr^{-1}	Feedstock	1,381,667	kg hr^{-1}

Table 13: Area 100 energy flows of feed handling

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Raw Feedstock	-134	MMkcal hr^{-1}	Raw Feedstock	-134	MMkcal hr^{-1}
	Water	-4,894	MMkcal hr^{-1}	Water (in)	-4,894	MMkcal hr^{-1}
OUT	Feedstock	-5,028	MMkcal hr^{-1}	Feedstock	-5,028	MMkcal hr^{-1}

5.1.3 Area 200: Mass Flows and energy flows for AD

Table 14 shows mass flow results and Table 15 shows energy flow results for the anaerobic digestion (plus cotreatment) area of the model. These mass flow values are the different for the two scenarios. The cotreatment scenario produced 28% more biogas per hour in comparison to the no cotreatment scenario. This result was expected since cotreatment enhances carbohydrate solubilization, resulting in a higher yield of biogas.

Table 14: Area 200 mass flows of AD

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Feedstock	1,381,667	kg hr^{-1}	Feedstock	1,381,667	kg hr^{-1}
OUT	Biogas	35,944	kg hr^{-1}	Biogas	45,847	kg hr^{-1}
	CO_2:CH_4 ratio	50.1:49.8	% molar	CO_2:CH_4 ratio	50.2:49.7	% molar
	Digestate	1,345,723	kg hr^{-1}	Digestate	1,335,820	kg hr^{-1}

Table 15: Area 200 energy flows of AD[2]

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Feedstock	-5,028	MMkcal hr^{-1}	Feedstock	-5,028	MMkcal hr^{-1}
OUT	Biogas	-71	MMkcal hr^{-1}	Biogas	-91	MMkcal hr^{-1}
	Digestate	-4,921	MMkcal hr^{-1}	Digestate	-4,903	MMkcal hr^{-1}

[2] Remaining differences in energy flows are attributed to heat duty of the reactor

5.1.4 Area 300: Mass and energy flows for upgrading

Table 16 shows results for the upgrading area of the model. The cotreatment scenario produces 27% more biomethane per hour in comparison to the no cotreatment scenario. Biomethane product quantities are reported at pipeline quality, 98% CH_4. This result was expected since the inflow of biogas is higher.

Table 16: Area 300 mass flows of upgrading

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Biogas	35,944	kg hr^{-1}	Biogas	45,847	kg hr^{-1}
	Digestate	1,345,723	kg hr^{-1}	Digestate	1,335,820	kg hr^{-1}
OUT	Liquid Digestate	1,300,204	kg hr^{-1}	Liquid Digestate	1,300,578	kg hr^{-1}
	Solid Digestate	45,518	kg hr^{-1}	Solid Digestate	35,241	kg hr^{-1}
	Water	3,567	kg hr^{-1}	Water	4,545	kg hr^{-1}
	Waste gas	23,945	kg hr^{-1}	Waste gas	30,576	kg hr^{-1}
	98% CH_4	8,431	kg hr^{-1}	98% CH_4	10,727	kg hr^{-1}

Table 17: Area 300 energy flows of upgrading[3]

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Biogas	-71	MMkcal hr^{-1}	Biogas	-91	MMkcal hr^{-1}
	Digestate	-4,921	MMkcal hr^{-1}	Digestate	-4,903	MMkcal hr^{-1}
OUT	Liquid Digestate	-4,906	MMkcal hr^{-1}	Liquid Digestate	-4,901	MMkcal hr^{-1}
	Solid Digestate	-55	MMkcal hr^{-1}	Solid Digestate	-40	MMkcal hr^{-1}
	Water	-14	MMkcal hr^{-1}	Water	-17	MMkcal hr^{-1}
	Waste gas	-51	MMkcal hr^{-1}	Waste gas	-65	MMkcal hr^{-1}
	98% CH_4	-10	MMkcal hr^{-1}	98% CH_4	-12	MMkcal hr^{-1}

5.1.5 Area 400: Mass flows for storage

Table 18 shows results for the storage area of the model. This is the overall output of the model. The 98% CH_4 stream is the mass flow of interest for the economic analysis. This will provide the project revenue. The cotreatment scenario produces 27% more biomethane than the no cotreatment scenario.

[3] Remaining differences in energy flows are attributed to heat duty of the pump, separators, compressors and cooler

Table 18: Area 400 mass flows for storage

	No cotreatment scenario			Cotreatment Scenario		
	Stream	Amount	Units	Stream	Amount	Units
IN	Liquid Digestate	1,300,204	kg hr^{-1}	Liquid Digestate	1,300,578	kg hr^{-1}
	Solid Digestate	45,518	kg hr^{-1}	Solid Digestate	35,241	kg hr^{-1}
	98% CH_4	8,431	kg hr^{-1}	98% CH_4	10,727	kg hr^{-1}

5.1.6 Summary of Process Mass Flow

Figure 11 shows the mass flows of each area for the cotreatment and no cotreatment scenario at a 2000 Mg day^{-1} scale. Mass is conserved throughout the process. In the feed handling area, mass flows of feedstock and water are equal for both scenarios (2000 Mg day^{-1} mixed with water to a 6% solid content). The anaerobic digestion area has different mass flows for each scenario. The cotreatment scenario produces 28% more biogas than the no cotreatment scenario. And finally, in the upgrading area, the cotreatment scenario produces 27% more biomethane than the no cotreatment scenario.

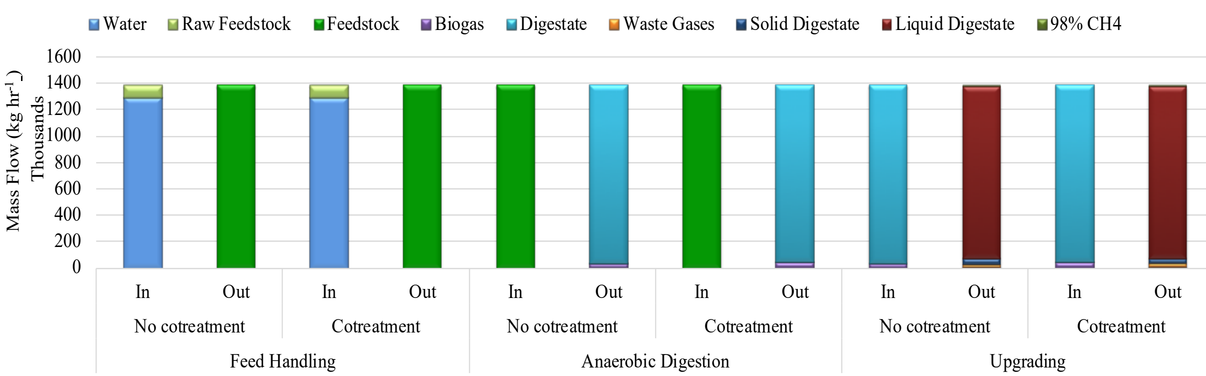


Figure 11: Process mass flows summarized by area

5.1.7 Sensitivity Analysis

The fractional carbohydrate extent of conversion factor of reactions 1, 4, 7 and 10 (Table 9) was varied for both scenarios. For the no cotreatment scenario, the carbohydrate solubilization was varied according to the standard deviation reported Liang et al. (2018). For the cotreatment scenario it was varied assuming a 5% solubilization increase as a minimum value and 30% solubilization increase as a maximum value. This assumption is based on previously reported solubilization rates of cotreatment assisted fermentation (Balch et al., 2017; Paye et al., 2016). minimum, average, and maximum value was estimated based on the varied percentage carbohydrate solubilization. Previous models have assumed conversion rates of 75%-90% (Humbird et al., 2011) without cotreatment, and previous cotreatment studies have reported up to 88% conversion with cotreatment (Balch et al., 2017).

Table 19 and Figure 12 present the results for the sensitivity analysis. The cotreatment scenario shows at least a 20% increase in biogas production or more compared to the no cotreatment scenario. This is a linear function , so for a 1% increase in carbohydrate solubilization, there is an increase of 131 kg hr-1 biomethane production at the 2000 Mg day-1 scale.

Table 19: Sensitivity analysis results for carbohydrate solubilization effect on biomethane mass flow rate

	No cotreatment scenario			Cotreatment scenario		
	Minimum	Average	Maximum	Minimum	Average	Maximum
Fractional carbohydrate conversion (%)	63.0	64.4	65.8	69.4	81.9	94.4
98% CH_4 flow (kg hr^{-1})	8,248	8,432	8,615	9,087	10,725	12,364
Energy Conversion Efficiency (%)	23.4	24.4	25.0	23.9	31.1	35.9

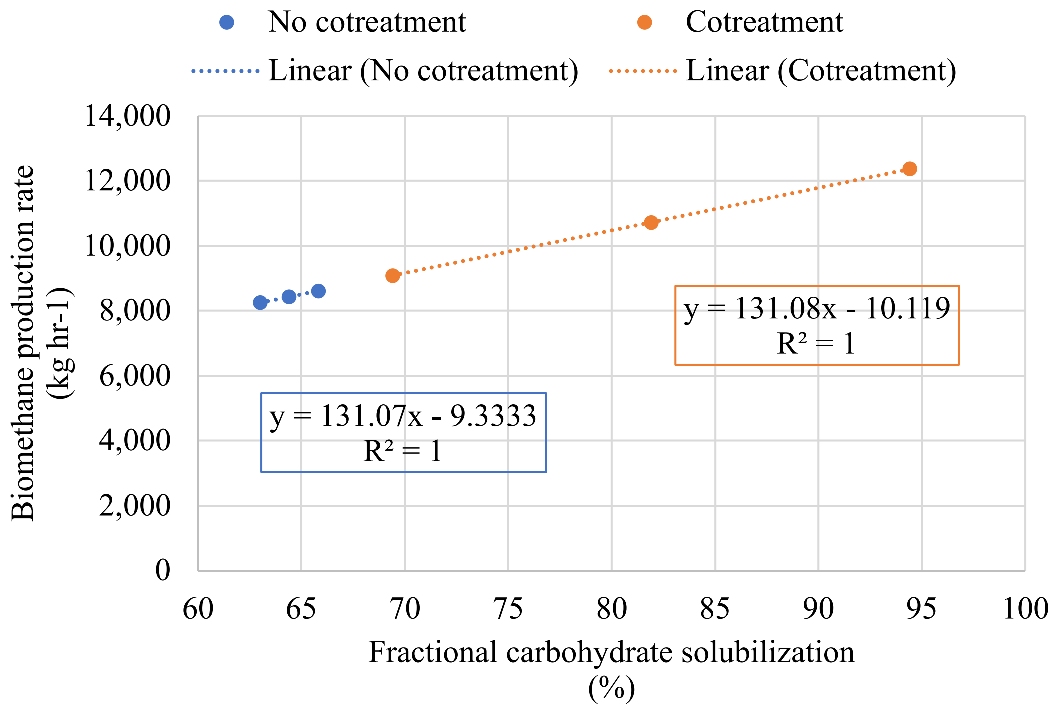


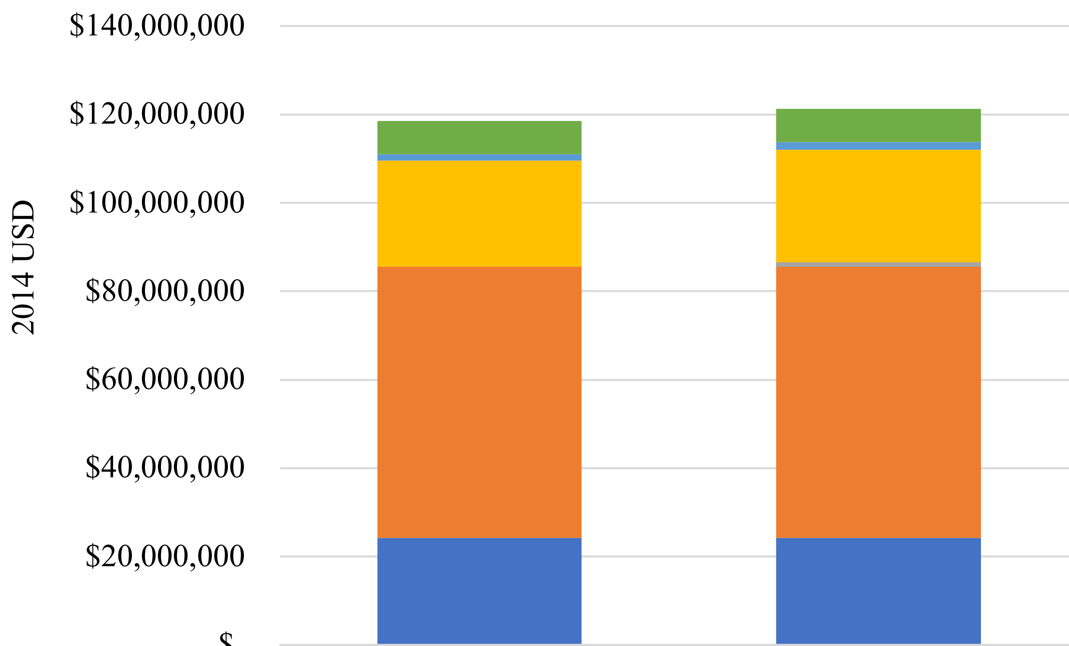
Figure 12: Sensitivity analysis results for carbohydrate solubilization rate effect on biomethane mass flow rate

5.2 Economic Analysis Phase

The following section presents results of the economic analysis phase. The capital expenses and operating expenses are described, and finally the minimum fuel selling price is summarized.

5.2.1 Capital Expenses

The total installed costs (TIC) of the project are shown on Figure 13. The cotreatment scenario has a higher TIC than the no cotreatment scenario. This is mainly due to the milling equipment in the anaerobic digestion area and the increase of biogas flow that goes through the upgrading area.



	No Cotreatment	Cotreatment
Utilities	$7,542,602	$7,542,602
Storage	$1,485,973	$1,681,050
Upgrading	$24,035,752	$25,512,704
Cotreatment	$0	$1,001,358
Anaerobic Digestion	$61,225,148	$61,225,148
Feed Handling	$24,261,817	$24,261,817

Figure 13: Total installed costs for switchgrass fed AD with and without cotreatment

Table 20 and Table 21 show a more detailed breakdown of the total capital investment for the no cotreatment and cotreatment scenarios respectively. Furthermore, Appendix B: Equipment Costs includes the detailed equipment, scaling factors and capital costs for each area. The total capital investment for the cotreatment scenario is 2.3% higher than for the no cotreatment scenario at the 2000 Mg day^{-1} scale.

Table 20: Total project cost for no cotreatment scenario of switchgrass fed AD

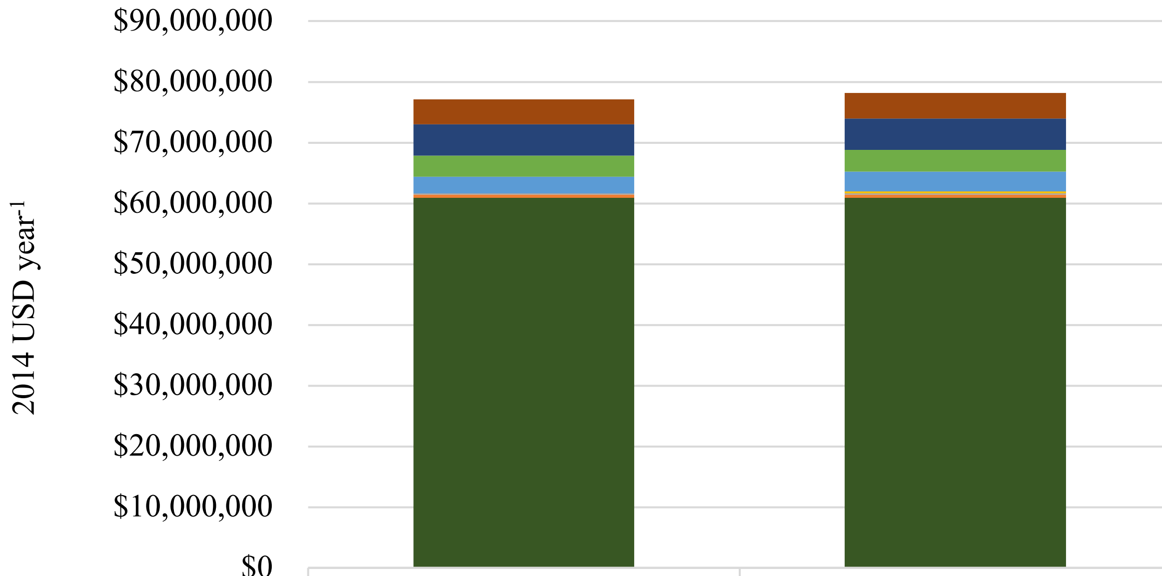
Total Capital Investment					
Process Area	**Description**		**Purchased Cost**	**Installed Cost**	**ISBL**
A100	Feed Handling		$14,271,657	$24,261,817	
A200	Anaerobic Digestion		$61,225,148	$61,225,148	$61,225,148
A300	Cotreatment		$0	$0	$0
A400	Upgrading		$23,920,559	$24,035,752	$24,035,752
A500	Storage		$990,648	$1,485,973	$1,485,973
A600	Utilities		$4,361,648	$7,542,602	
Subtotal			**$104,769,660**	**$118,551,291**	**$86,746,872**
Warehouse		4.0% of ISBL		$3,469,875	
Site Development		9.0% of ISBL		$7,807,219	
Additional Piping		4.5% of ISBL		$3,903,609	
Total Direct Costs (TDC)				**$133,731,993**	
Prorateable Expenses		10.0% of TDC		$13,373,199	
Field Expenses		10.0% of TDC		$13,373,199	
Home Office & Construction Fee		20.0% of TDC		$26,746,399	
Project Contingency		10.0% of TDC		$13,373,199	
Other Costs (Start-Up, Permits, etc.)		10.0% of TDC		$13,373,199	
Total Indirect Costs				**$80,239,196**	
Fixed Capital Investment (FCI)				**$213,971,190**	
Land				$1,848,000	
Working Capital		5.0% of FCI		$10,698,559	
Total Capital Investment (TCI)				**$226,517,749**	

Table 21: Total project cost for cotreatment scenario of switchgrass fed AD

Total Capital Investment				
Process Area	**Description**	**Purchased Cost**	**Installed Cost**	**ISBL**
A100	Feed Handling	$14,271,657	$24,261,817	
A200	Anaerobic Digestion	$61,225,148	$61,225,148	$61,225,148
A300	Cotreatment	$1,001,358	$1,001,358	$1,001,358
A400	Upgrading	$25,397,511	$25,512,704	$25,512,704
A500	Storage	$1,120,700	$1,681,050	$1,681,050
A600	Utilities	$4,361,648	$7,542,602	
Subtotal		**$107,378,023**	**$121,224,679**	**$89,420,261**
Warehouse		4.0% of ISBL	$3,576,810	
Site Development		9.0% of ISBL	$8,047,823	
Additional Piping		4.5% of ISBL	$4,023,912	
Total Direct Costs (TDC)			**$136,873,225**	
Prorateable Expenses		10.0% of TDC	$13,687,322	
Field Expenses		10.0% of TDC	$13,687,322	
Home Office & Construction Fee		20.0% of TDC	$27,374,645	
Project Contingency		10.0% of TDC	$13,687,322	
Other Costs (Start-Up, Permits, etc.)		10.0% of TDC	$13,687,322	
Total Indirect Costs			**$82,123,935**	
Fixed Capital Investment (FCI)			**$218,997,160**	
Land			$1,848,000	
Working Capital		5.0% of FCI	$10,949,858	
Total Capital Investment (TCI)			**$231,795,018**	

5.2.2 Variable and Fixed Operating Costs

This section describes the total operating costs of the project. These include variable and fixed costs. Variable operating costs include feedstock, raw materials needed and electricity. Fixed operating costs include salaries, maintenance and property insurance. Figure 14 summarizes the project operating costs. Appendix C: Operating Costs includes a breakdown of all the costs. The cotreatment scenario shows a 1.4% increase in total operating expenses 2000 Mg day^{-1} scale. This is mainly due to the added maintenance and electricity cost of the cotreatment system.



	No Cotreatment	Cotreatment
Overhead	$4,100,000	$4,240,000
Labor & Supervision	$5,170,000	$5,170,000
Utilities	$3,550,000	$3,550,000
Upgrading	$2,690,000	$3,230,000
Cotreatment	$0	$360,000
Anaerobic Digestion	$250,000	$250,000
Feed Handling	$460,000	$460,000
Feedstock	$60,930,000	$60,930,000

Figure 14: Total operating expenses per year of switchgrass fed AD with and without cotreatment

5.2.3 Sensitivity Analysis

Finally, the minimum fuel selling price for each scenario was calculated. The previous mass flow sensitivity analysis was considered for the minimum, average and maximum production of biomethane. The average MFSP (per gallon gasoline equivalent) for the cotreatment scenario is estimated to be $0.86 lower than the cotreatment (Table 22). The DOE goal is to generate biofuels with a modeled cost of \$3 GGE^{-1} or less (U.S. Department of Energy, 2016a). The cotreatment scenario meets such requirement at the maximum production simulated, 2000 Mg dry biomass day^{-1}, with the baseline assumption of a 30% increase in carbohydrate solubilization in comparison to no cotreatment).

Table 22: Minimum fuel selling price, accounting for sensitivity analysis

	No cotreatment scenario			Cotreatment scenario		
Fractional carbohydrate solubilization (%)	63.0	64.4	65.8	69.4	81.9	94.4
Minimum Fuel Selling Price (2014 USD GGE^{-1})	4.31	4.23	4.15	3.96	3.37	2.82

A sensitivity analysis was performed to determine the effect of carbohydrate solubilization on the MFSP (Figure 15). The relationship is linear and so, a 1% increase in carbohydrate solubilization decreases the price of switchgrass fed AD without cotreatment by \$0.06 GGE^{-1}. For the cotreatment assisted AD scenario, the MFSP decreases by \$0.05 GGE^{-1} per 1% increase in carbohydrate solubilization. This difference in slope and intercept is attributed to the added costs of milling equipment in the cotreatment scenario.

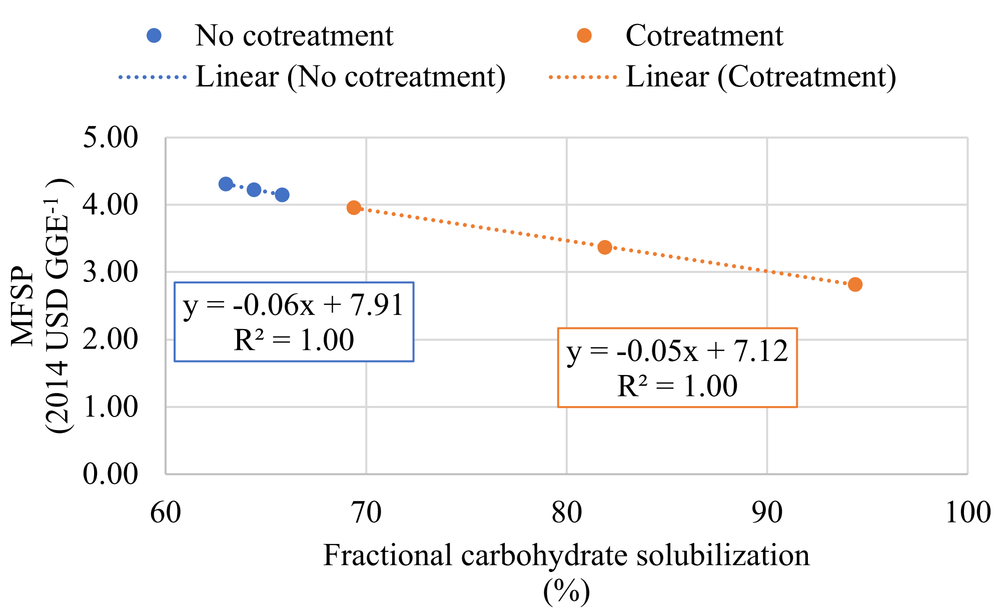


Figure 15: Sensitivity analysis of carbohydrate solubilization effect on MFSP

The MFSP was also calculated for different scales of AD with and without cotreatment (Figure 16), at average carbohydrate solubilization, to examine the effect of economies of scale on MFSP. The two horizontal lines indicate reference prices, including a 2016 average market price of CNG of \$2.09 GGE^{-1} (U.S. Department of Energy, 2016b). In addition, the renewable natural gas market price is supplemented by additional RIN credit that averaged \$3.36 GGE^{-1} CH_4 in 2016 (EPA, 2016), for a total of \$5.45 GGE^{-1} of RNG (2016 USD). At none of the scales or scenarios is renewable CNG production from switchgrass through AD economically competitive with fossil natural gas without policy incentives. However, taking into account the renewable fuel standard incentives, switchgrass fed AD for renewable CNG production becomes economically feasible. Furthermore, cotreatment allows switchgrass fed AD to be economically feasible (including RINs) at lower scales than without cotreatment.

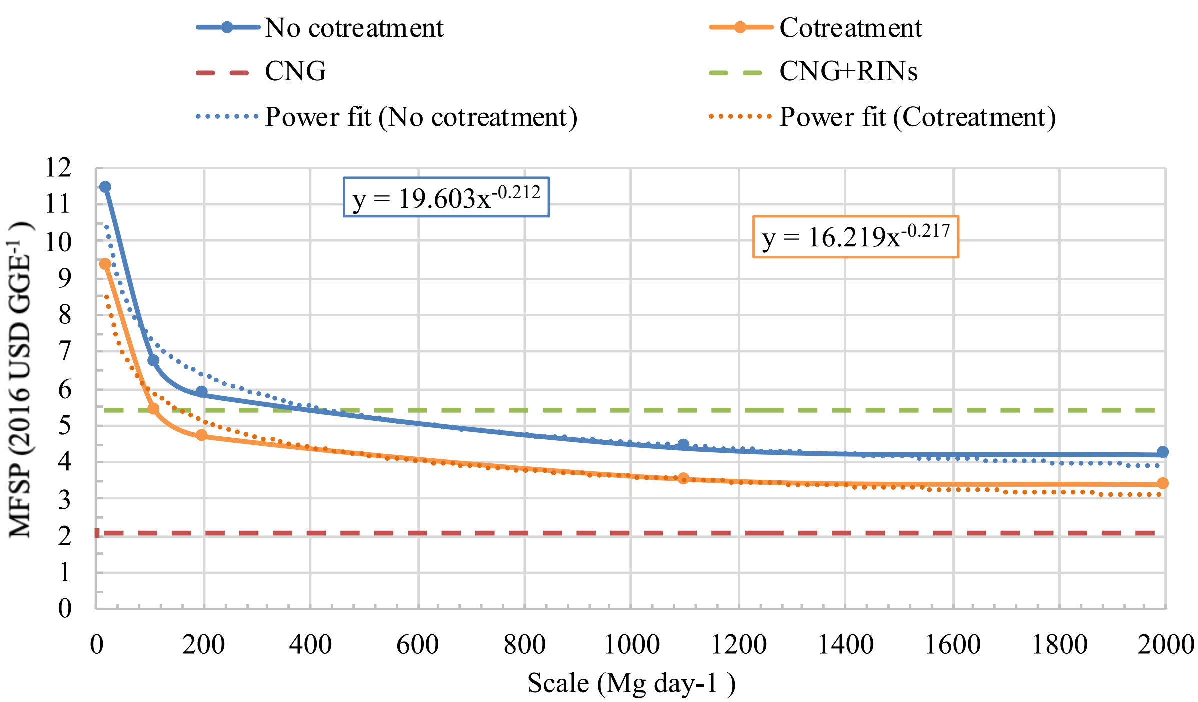


Figure 16: Scale effect on MFSP of switchgrass fed AD with and without cotreatment compared to CNG with (green dashed line) and without (red dashed line) RIN prices (2016 USD)

Although the power fit is a good representation at large scales, it is not representative of the behavior at smaller scales. Figure 17 shows a close up of these smaller scales fitted with a second order polynomial curve. This fit is better suited to describe MFSP behavior at smaller scales, especially less than 200 Mg dry biomass day^{-1}. When solving for a MFSP of \$5.45, the equations estimate scales greater than 134 Mg day^{-1} (no cotreatment) and 110 Mg day^{-1} (cotreatment) are economically feasible. However, because even these smaller scale regression equations deviate from the actual solution, these intercepts with the two price lines were solved numerically using the developed TEA model. According to this numerical solution, switchgrass fed AD becomes feasible at scales greater than 280 Mg day^{-1} (no cotreatment) and 104 Mg day^{-1} (cotreatment). This suggests that cotreatment allows for economically feasible operation of 2.7x smaller scale of switchgrass fed AD in comparison to no cotreatment.

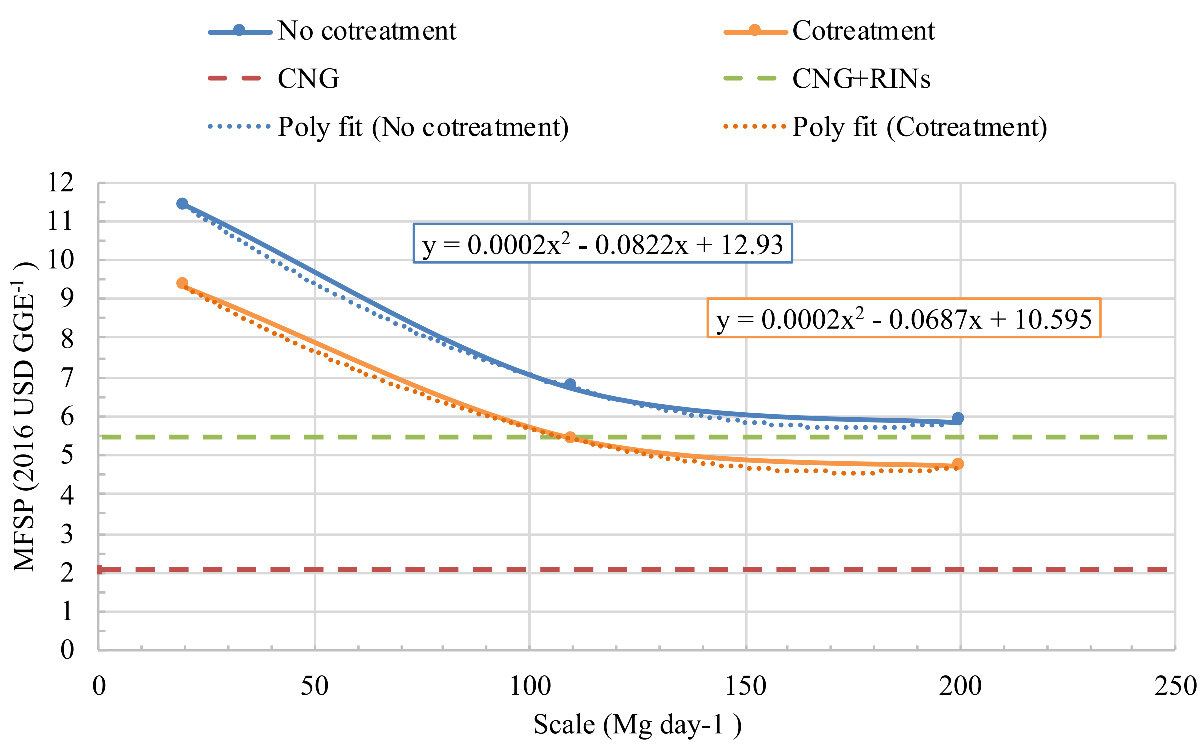


Figure 17: Sensitivity analysis of scale effect on MFSP focused on small scale

5.2.4 Economic Validation

Figure 18 shows a comparison of minimum fuel selling price (2016 USD GGE^{-1}) obtained of the model with previously reported values of cellulosic biofuel conversion technologies (Gnansounou and Dauriat, 2011; Huang et al., 2009; Humbird et al., 2011; Kazi et al., 2010). All estimated values are within the range of previously reported numbers although cotreatment assisted AD is on the least expensive extreme. These studies include other feedstocks and price assumptions and are presented to provide background on the price scale of cellulosic biofuels.

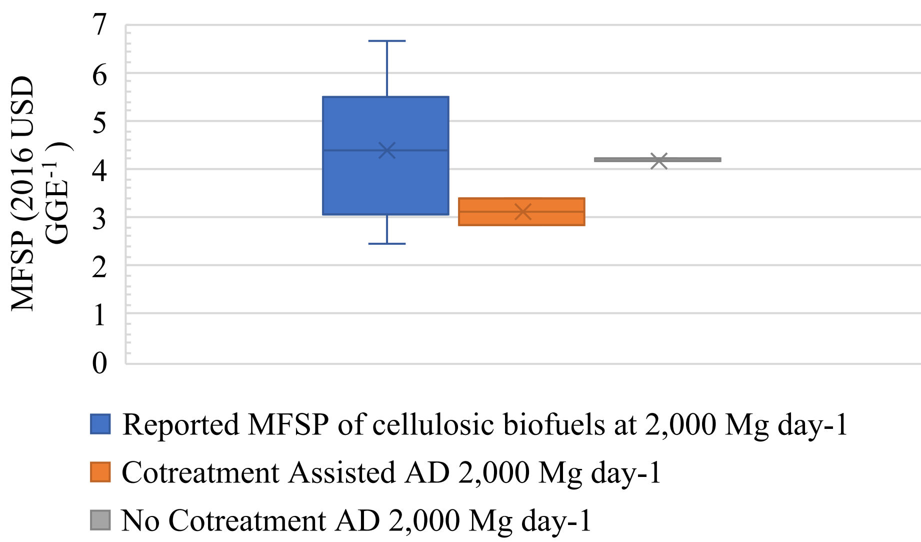


Figure 18: Comparison of economic analysis results with previously reported values for cellulosic biofuels at a 2,000 Mg day^{-1} (dry basis)

Chapter 6: Conclusions and Recommendations

Cotreatment research has previously indicated a potential economic benefit for pure culture, biomass-fed fermentation systems for ethanol production. The goal of this study was to examine potential economic benefits of cotreatment in mixed culture, biomass-fed anaerobic digestion systems for biomethane production.

Mass flow results show a potential increase of 27% more biomethane production with cotreatment in comparison to no cotreatment for a fixed ten day residence time at a 2000 Mg day^{-1} scale. The capital and operating expenses are higher for cotreatment, but because of this greater yield there is still a significant decrease in minimum fuel selling price when cotreatment is integrated in the process. The cost for cotreatment aided, mixed culture, biomass-fed anaerobic digestion systems for biomethane production is \$3.37 GGE^{-1} compared to \$4.23 GGE^{-1} for no cotreatment (average production scenario at 2000 Mg biomass day^{-1}). Under these conditions cotreatment decreases the MFSP by \$0.86, and the carbohydrate solubilization sensitivity analysis estimates a 5 cent reduction in MFSP per 1% increase in solubilization. At a 94.4% carbohydrate solubilization factor at the 2000 Mg day^{-1} scale, cotreatment aided switchgrass fed AD becomes cost competitive relative to CNG. However, when considering a RIN incentive of \$3.36 GGE^{-1} of CNG, switchgrass fed AD becomes economically feasible at scales greater than 280 Mg day^{-1} scale without cotreatment and at a 104 Mg day^{-1} with cotreatment.

Some limitations of the present work are limited information and uncertainties in milling energy requirements and scale up costs, uncertainty in experimental biogas production yield, and carbohydrate solubilization in cotreatment aided AD. These assumptions are based on cotreatment experimental result in pure culture fermentations. Future work that could extend this research includes: developing scenarios of anaerobic digestion of biomass plus manure mixes, more extensive milling energy tests to document energy requirements, and generating and integrating experimental results for fractional carbohydrate solubilization and biogas production rate of mixed culture systems. In addition, exploring cotreatment of mixed cultures at different solids contents, temperatures or retention times could further inform on favorable operating parameters of AD.

These findings could help with the planning of large-scale anaerobic digesters to reach government renewable energy policy standards, reduce greenhouse gas emissions and provide a sustainable, cost-efficient bioenergy resource.

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Appendix A: Aspen Plus Properties

Table A 1: Aspen Plus property set used. (Source: Humbird et al., 2011)

Component	Property	Quantity	Units	Reference
H2O	-	-	-	Native Aspen component
ETHANOL	-	-	-	Native Aspen component
GLUCOSE	-	-	-	Native Aspen component (dextrose)
GALACTOS	-	-	-	Duplicate of GLUCOSE
MANNOSE	-	-	-	Duplicate of GLUCOSE
XYLOSE	DHFORM	-216752.65	cal/mol	Native Aspen component (d-xylose) with DHFORM specified (5/6 of GLUCOSE DHFORM)
ARABINOS	-	-	-	Duplicate of XYLOSE
CELLOB	-	-	-	Cellobiose. Used native Aspen component sucrose
SUCROSE	-	-	-	Native Aspen component
GLUCOLIG	MW	162.1424		Glucose oligomers. Most properties from GLUCOSE; MW is GLUCOSE minus H2O
	DHFORM	-192875.34	cal/mol	Back-calculated to match ΔH_c of CELLULOS
GALAOLIG	-	-	-	Galactose oligomers. Duplicate of GLUCOLIG
MANOLIG	-	-	-	Mannose oligomers. Duplicate of GLUCOLIG
XYLOLIG	MW	132.11612		Xylose oligomers. Most properties from XYLOSE; MW is XYLOSE minus H2O
	DHFORM	-149412.58	cal/mol	Back-calculated to match ΔH_c of XYLAN
ARABOLIG	-	-	-	Arabinose oligomers. Duplicate of XYLOLIG
EXTRACT	-	-	-	Organic extractives. Duplicate of GLUCOSE
LGNSOL	-	-	-	Solubilized lignin. Native Aspen component vanillin (see note at LIGNIN)

Component	Property	Quantity	Units	Reference
HMF	MW	126.11		(5-hydroxymethylfurfural) Properties for HMF were estimated within Aspen using NIST TDE routines. Specify molecular structure, MW, TB and DHFORM. From [90].
	TB	532.7	K	
	DHFORM	-79774.53	cal/mol	
	DHVLWT-1	80550000	J/kmol	
	TC	731.012	K	NIST TDE
	PC	5235810	Pa	
	OMEGA	0.993646710		
	VC	0.3425	m^3/kmol	
	RKTZRA	0.198177974		
FURFURAL	-	-	-	Native Aspen component
AACID	-	-	-	Native Aspen component (acetic acid)
LACID	-	-	-	Native Aspen component (lactic acid)
XYLITOL	-	-	-	Native Aspen component
GLYCEROL	-	-	-	Native Aspen component
SUCCACID	-	-	-	Native Aspen component (Succinic acid)
NH3	-	-	-	Native Aspen component
H2SO4	-	-	-	Native Aspen component
NH4SO4	-	-	-	Native Aspen component (ammonium sulfate)
NH4ACET	PLXANT/1	-1E20	atm	Native Aspen component (ammonium acetate) forced non-volatile
DAP	-	-	-	Native Aspen component (diammonium phosphate)
HNO3	-	-	-	Native Aspen component
NANO3	-	-	-	Native Aspen component
NAOH	-	-	-	Native Aspen component
CNUTR	-	-	-	Cellulase nutrient mix. Duplicate of glucose
WNUTR	-	-	-	WWT nutrient mix. Duplicate of glucose
DENAT	-	-	-	Denaturant. Native Aspen component (n-heptane)
OIL	-	-	-	Corn oil antifoam. Native Aspen component (oleic acid)
O2	-	-	-	Native Aspen component

Component	Property	Quantity	Units	Reference
N2	-	-	-	Native Aspen component
NO	-	-	-	Native Aspen component
NO2	-	-	-	Native Aspen component
CO	-	-	-	Native Aspen component
CO2	-	-	-	Native Aspen component
CH4	-	-	-	Native Aspen component
H2S	-	-	-	Native Aspen component
SO2	-	-	-	Native Aspen component
CELLULOS	DHSFRM	-233200.06	cal/mol	Native Aspen component with specified heat of formation; back-calculated
GALACTAN	-	-	-	Duplicate of CELLULOS
MANNAN	-	-	-	Duplicate of CELLULOS
XYLAN	Formula	$C_5H_8O_4$ (monomer)		
	MW	132.117		
	DHSFRM	-182099.93	cal/mol	From [22]. Assumes the ratio of ΔH_c of glucose:xylose is the same for cellulose:xylan.
ARABINAN	-	-	-	Duplicate of xylan
LIGNIN	-	-	-	Used native Aspen component vanillin ($C_8H_8O_3$). The HHV of this compound (-23,906 BTU/kg) is very close to what we previously assumed for lignin as a custom component (-24,206)
ACETATE	-	-	-	Used native Aspen component acetic acid
PROTEIN	Formula	$CH_{1.57}O_{0.31}N_{0.29}S_{0.007}$		Wheat gliadin [91]
	MW	22.8396		
	DHSFRM	-17618	cal/mol	From literature value of gliadin ΔH_c [92]
	PLXANT/1	-1E20	atm	Forces non-volatility
ASH	-	-	-	Native Aspen component CaO
ENZYME	Formula	$CH_{1.59}O_{0.42}N_{0.24}S_{0.01}$		Provided by Novozymes [55]
	MW	24.0156		
	DHSFRM	-17618	cal/mol	copied from PROTEIN
DENZ	-	-	-	Denatured enzyme. Duplicate of enzyme

Component	Property	Quantity	Units	Reference
ZYMO	Formula	$CH_{1.8}O_{0.5}N_{0.2}$		Z. mobilis cell mass. Average composition of several microorganisms outlined in [93]
	MW	24.6264		
	DHSFRM	-31169.39	cal/mol	From [22] after [94]
TRICHO	Formula	$CH_{1.645}O_{0.445}N_{0.205}S_{0.005}$		T. reesei cell mass. Average of generic cell mass [56] and the ENZYME composition
	MW	23.8204		
	DHSFRM	-23200.01	cal/mol	Copied BIOMASS
BIOMASS	Formula	$CH_{1.64}O_{0.39}N_{0.23}S_{0.0035}$		WWT sludge.
	MW	23.238		
	DHSFRM	-23200.01	cal/mol	From [22] after [94]
TAR	-	-	-	Modeled as solid xylose
LIME	-	-	-	Aspen native component (calcium hydroxide)
CASO4	-	-	-	Aspen native component
C	-	-	-	Aspen native component (graphite carbon)

Table A 2: List of combustion energies (Low heating value and high heating values) used (Source: Humbird et al., 2011)

	MW	ΔHf (IG)	ΔHf (L)	ΔHf (S)																			LHV			HHV		
		cal/mol	cal/mol	cal/mol																			cal/mol	Gcal/kmol	BTU/lb	cal/mol	Gcal/kmol	BTU/lb
GLUCOSE	180.16	-260103	-300428		GLUCOSE	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723
XYLOSE	150.13	-216753	-249440		XYLOSE	+	5	O2	→	5	H2O	+	5	CO2									-509283	-0.509283	-6106	-561662	-0.561662	-6734
SUCROSE	342.30	-459301	-480900		SUCROSE	+	12	O2	→	11	H2O	+	12	CO2									-1282278	-1.282278	-6743	-1397513	-1.397513	-7349
EXTRACT	180.16	-260103	-300428		EXTRACT	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723
ARABINOS	150.13	-216753	-249440		ARABINOS	+	5	O2	→	5	H2O	+	5	CO2									-509283	-0.509283	-6106	-561662	-0.561662	-6734
GALACTOS	180.16	-260103	-300428		GALACTOS	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723
MANNOSE	180.16	-260103	-300428		MANNOSE	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723
GLUCOLIG	162.14	-192875	-233200		GLUCOLIG	+	6	O2	→	5	H2O	+	6	CO2									-619511	-0.619511	-6877	-671890	-0.671890	-7459
CELLOB	342.30	-459301	-480900		CELLOB	+	12	O2	→	11	H2O	+	12	CO2									-1282278	-1.282278	-6743	-1397513	-1.397513	-7349
XYLOLIG	132.12	-149413	-182100		XYLOLIG	+	5	O2	→	4	H2O	+	5	CO2									-518866	-0.518866	-7069	-560770	-0.560770	-7640
MANOLIG	162.14	-192875	-233200		MANOLIG	+	6	O2	→	5	H2O	+	6	CO2									-619511	-0.619511	-6877	-671890	-0.671890	-7459
GALAOLIG	162.14	-192875	-233200		GALAOLIG	+	6	O2	→	5	H2O	+	6	CO2									-619511	-0.619511	-6877	-671890	-0.671890	-7459
ARABOLIG	132.12	-149413	-182100		ARABOLIG	+	5	O2	→	4	H2O	+	5	CO2									-518866	-0.518866	-7069	-560770	-0.560770	-7640
XYLITOL	152.15	-222246	-243145	-267149	XYLITOL	+	5.5	O2	→	6	H2O	+	5	CO2									-573334	-0.573334	-6783	-636190	-0.636190	-7527
LGNSOL	152.15	-88134	-108248		LGNSOL	+	8.5	O2	→	4	H2O	+	8	CO2									-874683	-0.874683	-10348	-916587	-0.916587	-10844
ETHANOL	46.07	-56117	-66278		ETHANOL	+	3	O2	→	3	H2O	+	2	CO2									-294968	-0.294968	-11525	-326395	-0.326395	-12753
H2O	18.02	-57756	-68232	-69963	NA																							
FURFURAL	96.09	-36066	-47738		FURFURAL	+	5	O2	→	2	H2O	+	5	CO2									-537716	-0.537716	-10073	-558667	-0.558667	-10466
HMF	126.11	-79775	-99677		HMF	+	6	O2	→	3	H2O	+	6	CO2									-637521	-0.637521	-9099	-668949	-0.668949	-9548
H2SO4	98.08	-175600	-189490		H2SO4	+	-0.5	O2	→	1	H2O							+	1	SO2			60835	0.060835	1116	50359	0.050359	924
N2	28.01	0	--	--	NA																							
CO2	44.01	-93988	-95245		NA																							
O2	32.00	0	--	--	NA																							
CH4	16.04	-17799	--	--	CH4	+	2	O2	→	2	H2O	+	1	CO2									-191702	-0.191702	-21509	-212654	-0.212654	-23860
NO	30.01	21556	--		NO	+	0.5	O2	→												1	NO2	-13631	-0.013631	-818	-13631	-0.013631	-818
NO2	46.01	7925	-1179		NA																							
NH3	17.03	-10963	-15706		NH3	+	0.75	O2	→	1.5	H2O				+	0.5	N2						-75672	-0.075672	-7998	-91386	-0.091386	-9659
LACID	90.08	-143212	-163122		LACID	+	3	O2	→	3	H2O	+	3	CO2									-292112	-0.292112	-5837	-323539	-0.323539	-6465
AACID	60.05	-103373	-108966		AACID	+	2	O2	→	2	H2O	+	2	CO2									-194523	-0.194523	-5831	-215474	-0.215474	-6459
NH4SO4	132.14	-282029	-288994	-282041	NH4SO4	+	1	O2	→	4	H2O				+	1	N2	+	1	SO2			-12930	-0.012930	-176	-54834	-0.054834	-747
NH4ACET	77.08	-147163	-154701	-147163	NH4ACET	+	2.75	O2	→	3.5	H2O	+	2	CO2	+	0.5	N2						-235422	-0.235422	-5498	-272088	-0.272088	-6354
GLYCEROL	92.09	-138029	-159483		GLYCEROL	+	3.5	O2	→	4	H2O	+	3	CO2									-353507	-0.353507	-6909	-395411	-0.395411	-7728
SUCCACID	118.09	-196546	-221160	-231172	SUCCACID	+	3.5	O2	→	3	H2O	+	4	CO2									-328062	-0.328062	-5001	-359490	-0.359490	-5480
DAP	132.06	-276457	-283996	-374250	NA																							
DENAT	100.20	-44819	-53568		DENAT	+	11	O2	→	8	H2O	+	7	CO2									-1066400	-1.066400	-19156	-1150207	-1.150207	-20662
OIL	282.47	-169628	-196067		OIL	+	25.5	O2	→	17	H2O	+	18	CO2									-2477579	-2.477579	-15788	-2655669	-2.655669	-16923
CNUTR	180.16	-260103	-300428		CNUTR	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723
WNUTR	180.16	-260103	-300428		WNUTR	+	6	O2	→	6	H2O	+	6	CO2									-610039	-0.610039	-6095	-672895	-0.672895	-6723

	MW	ΔHf (IG)	ΔHf (L)	ΔHf (S)																					LHV			HHV	
		cal/mol	cal/mol	cal/mol																				cal/mol	Gcal/kmol	BTU/lb	cal/mol	Gcal/kmol	BTU/lb
SO2	64.06	-70899	-76380		NA																								
H2S	34.08	-4927	-8300		H2S	+	1.5	O2	→	1	H2O							+	1	SO2				-123728	-0.123728	-6535	-134204	-0.134204	-7088
CO	28.01	-26400	--		CO	+	0.5	O2	→				1	CO2										-67589	-0.067589	-4343	-67589	-0.067589	-4343
HNO3	63.01	-32077	-41406		HNO3	+	-1.75	O2	→	0.5	H2O				+	0.5	N2							12528	0.012528	358	-2039	-0.002039	-58
NA2O			-6965		NA																								
NAOH	40.00	-47234	-67046		NAOH	+	0	O2	→	0.5	H2O										+	0.5	NA2O	34685	0.034685	1561	9636	0.009636	434
NANO3	84.99		-118756		NANO3	+	-1.25	O2	→							0.5	N2				+	0.5	NA2O	115273	0.115273	2441	115273	0.115273	2441
CELLULOS	162.14			-233200	CELLULOS	+	6	O2	→	5	H2O	+	6	CO2										-619511	-0.619511	-6877	-671890	-0.671890	-7459
XYLAN	132.12			-182100	XYLAN	+	5	O2	→	4	H2O	+	5	CO2										-518866	-0.518866	-7069	-560770	-0.560770	-7640
LIGNIN	152.15			-108248	LIGNIN	+	8.5	O2	→	4	H2O	+	8	CO2										-874683	-0.874683	-10348	-916587	-0.916587	-10844
ENZYME	24.02			-17900	ENZYME	+	1.1975	O2	→	0.795	H2O	+	1	CO2	+	0.12	N2	+	0.01	SO2				-122713	-0.122713	-9198	-130333	-0.130333	-9769
DENZ	24.02			-17900	DENZ	+	1.1975	O2	→	0.795	H2O	+	1	CO2	+	0.12	N2	+	0.01	SO2				-122713	-0.122713	-9198	-130333	-0.130333	-9769
BIOMASS	23.24			-23200	BIOMASS	+	1.2185	O2	→	0.82	H2O	+	1	CO2	+	0.115	N2	+	0.0035	SO2				-118397	-0.118397	-9171	-126739	-0.126739	-9817
ZYMO	24.63			-31169	ZYMO	+	1.2	O2	→	0.9	H2O	+	1	CO2	+	0.1	N2							-114800	-0.114800	-8391	-124228	-0.124228	-9080
ACETATE	60.05	-103373		-108992	ACETATE	+	2	O2	→	2	H2O	+	2	CO2										-194497	-0.194497	-5830	-215449	-0.215449	-6458
ARABINAN	132.12			-182100	ARABINAN	+	5	O2	→	4	H2O	+	5	CO2										-518866	-0.518866	-7069	-560770	-0.560770	-7640
MANNAN	162.14			-233200	MANNAN	+	6	O2	→	5	H2O	+	6	CO2										-619511	-0.619511	-6877	-671890	-0.671890	-7459
GALACTAN	162.14			-233200	GALACTAN	+	6	O2	→	5	H2O	+	6	CO2										-619511	-0.619511	-6877	-671890	-0.671890	-7459
TAR	150.13			-182100	TAR	+	5	O2	→	5	H2O	+	5	CO2										-576623	-0.576623	-6913	-629002	-0.629002	-7541
ASH	56.08	10485		-151688	NA																								
TRICHO	23.82			-23200	TRICHO	+	1.19375	O2	→	0.8225	H2O	+	1	CO2	+	0.1025	N2	+	0.005	SO2				-118647	-0.118647	-8966	-126909	-0.126909	-9590
PROTEIN	22.84			-17618	PROTEIN	+	1.2445	O2	→	0.785	H2O	+	1	CO2	+	0.145	N2	+	0.007	SO2				-122205	-0.122205	-9631	-129932	-0.129932	-10240
CARBON	12.01			0	C	+	1	O2	→				1	CO2										-93988	-0.093988	-14085	-93988	-0.093988	-14085
LIME	74.09	-145878		-235522	Ca(OH)2				→	1	H2O										+	1	Ash	26078	0.026078	634	15602	0.015602	379
CASO4	136.14	-315682		-342531	CaSO4	+	-0.5	O2	→										1	SO2	+	1	Ash	119944	0.119944	1586	119944	0.119944	1586

Appendix B: Equipment Costs

Table B 1: Detailed capital costs for switchgrass fed AD with cotreatment

Total Capital Costs													Total:		$107,378,023	$121,224,679
			Quote details			Unit Sizing Details										
PATH	Config	EQUIPMENT TITLE	QUOTE INFO	$ (individual quote)	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
A100		Feed Handling														
0																
		Transfer Conveyor	Humbird 2011	$5,397,000	2009	$5,397,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$5,234,753	$5,778,388.93	$9,823,261
		High Angle Transfer Conveyor	Humbird 2011	INCLUDED												
		Reversing Load-in Conveyor	Humbird 2011	INCLUDED												
		Dome Reclaim System	Humbird 2011	$3,046,000	2009	$3,046,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$2,954,430	$3,261,251.19	$5,544,127
		Reclaim Conveyor	Humbird 2011	INCLUDED												
		High Angle Transfer Conveyor	Humbird 2011	INCLUDED												
		Elevated Transfer Conveyor	Humbird 2011	INCLUDED												
		Process Feed Conveyor	Humbird 2011	INCLUDED												
		Truck Scale	Humbird 2011	$110,000	2009	$110,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$106,693	$117,773.35	$200,215
		Truck Dumper	Humbird 2011	$484,000	2009	$484,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$469,450	$518,202.75	$880,945
		Truck Dumper Hopper	Humbird 2011	$502,000	2009	$502,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$486,909	$537,474.75	$913,707
		Concrete Feedstock Storage Dome	Humbird 2011	$3,500,000	2009	$3,500,000	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$3,394,781	$3,747,333.93	$6,370,468
		Belt Scale	Humbird 2011	$10,790	2009	$10,790	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$10,466	$11,552.50	$19,639
		Dust Collection System	Humbird 2011	$279,900	2009	$279,900	STRM.A100.102	94697	kg/hr	0.60	1.7	90000	0.95	$271,486	$299,679.65	$509,455
													Subtotals:	$12,928,967	$14,271,657	$24,261,817
A200		Anaerobic Digestion														
0																
		Biogas Blower	Humbird 2011	INCLUDED												
		Biogas Emergency Flare	Humbird 2011	$32,955	2010	$32,955	STRM.A100.103	393100	kg/hr	0.60	1.00	1381666.7	3.51479692	$70,058.60	$73,277	$73,277
		Anaerobic Reactor Feed Pump	Humbird 2011	$231,488	2010	$231,488	STRM.A100.103	393100	kg/hr	0.60	1.00	1381666.7	3.51479692	$492,117.29	$514,722	$514,722
		Anaerobic Reactor Recirc Pump	Humbird 2011	INCLUDED												
		Waste Anaerobic Sludge Pump	Humbird 2011	$93,300	2010	$93,300	STRM.A100.103	393100	kg/hr	0.60	1.00	1381666.7	3.51479692	$198,345.24	$207,456	$207,456
		Return Activated Sludge Pump	Humbird 2011	$177,300	2010	$177,300	STRM.A100.103	393100	kg/hr	0.60	1.00	1381666.7	3.51479692	$376,919.74	$394,233	$394,233
		Anaerobic Basin	Humbird 2011	$27,000,000	2010	$27,000,000	STRM.A100.103	393100	kg/hr	0.60	1.00	1381666.7	3.51479692	$57,398,944.54	$60,035,461	$60,035,461
													Subtotals:	$58,536,385.41	$61,225,148	$61,225,148
A300		Cotreatment														
0																
		Colloid Mill	IKA, Personal communication	$250,000	2017	$250,000	STRM.A100.103	80000	L/hr	0.61	1.00	759157.51	9.48946889	$986,410.24	$1,001,358	$1,001,358
													Subtotals:	$986,410.24	$1,001,358	$1,001,358
A400		Upgrading														
0																
		Upgrading Equipment	Bauer et al. 2013	$1,581	2012	$2,301,936	STRM.A300.308	1400	m3/hr	0.50	1.00	30821.943	22.0156737	$10,800,882.34	$10,964,561	$10,964,561
		Membrane replacement	Bauer et al. 2013	INCLUDED												
		Waste Vapor Condenser	Humbird 2011	$34,000	2009	$34,000	Heat.A200.QH244	2	Gcal/hr	0.70	2.2	7	3.83	$86,962	$95,993.62	$211,186
		Anerobic Sludge Screw	Humbird 2011	$25,000	2010	$25,000	STRM.A200.203	393100	kg/hr	0.60	1.0	1335820	3.40	$52,082	$54,474.20	$54,474
		Centrifuge Feed Pump	Humbird 2011	$61,200	2010	$61,200	STRM.A200.203	393100	kg/hr	0.60	1.0	1335820	3.40	$127,497	$133,352.85	$133,353
		Centrifuge	Humbird 2011	$6,493,500	2010	$6,493,500	STRM.A200.203	393100	kg/hr	0.60	1.0	1335820	3.40	$13,527,757	$14,149,129.87	$14,149,130
													Subtotals:	$24,595,180.11	$25,397,511	$25,512,704
A500		Storage														
0																
		Sludge Storage Tank	W. Fuchs and B. Drosg, (2013)	$7	2013	$7	VOLLIQDIG	1	m3	1.00	1.5	31280	21722	$149,881.80	$152,206.78	$228,310
		Gas Storage Tank	Smith, M., Gonzales, J., 2014 DOE	$1,200,000	2014	$1,200,000	VOL.CH4	1558	m3	0.60	1.50	1090	0.69961489	$968,493.42	$968,493.42	$1,452,740
													Subtotals:	$ 1,118,375.22	$1,120,700	$1,681,050
A600		Utilities														
0																
		Cooling Tower System	Humbird 2011	$1,375,000	2010	$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	11922359	1.19	$1,524,531	$1,594,558	$2,391,837
		Plant Air Compressor	Humbird 2011	$28,000	2010	$28,000	DRY101	83333	kg/hr	0.60	1.6	83333	1.00	$28,000	$29,286	$46,858
		Chilled Water Package	Humbird 2011	$1,275,750	2010	$1,275,750	heat.a900.qchwop	14	Gcal/hr	0.60	1.6	13	0.95	$1,234,569	$1,291,276	$2,066,042
		CIP System	Humbird 2011	$421,000	2009	$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	$694,222	$766,318	$1,379,373
		Cooling Water Pump	Humbird 2011	$283,671	2010	$283,671	strm.a900.945	10982556	kg/hr	0.80	3.1	11922359	1.09	$302,930	$316,844	$982,217
		Make-up Water Pump	Humbird 2011	$6,864	2010	$6,864	strm.a900.904	155564	kg/hr	0.80	3.1	147136	0.95	$6,565	$6,866	$21,286
		Process Water Circulating Pump	Humbird 2011	$15,292	2010	$15,292	strm.a900.905	518924	kg/hr	0.80	3.1	523776	1.01	$15,406	$16,114	$49,953
		Instrument Air Dryer	Humbird 2011	$15,000	2009	$15,000	DRY101	83333	kg/hr	0.60	1.8	83333	1.00	$15,000	$16,558	$29,804
		Plant Air Receiver	Humbird 2011	$16,000	2009	$16,000	DRY101	83333	kg/hr	0.60	3.1	83333	1.00	$16,000	$17,662	$54,751
		Process Water Tank No. 1	Humbird 2011	$250,000	2009	$250,000	strm.a900.905	451555	kg/hr	0.70	1.7	523776	1.16	$277,361	$306,165	$520,480
													Subtotals:		$4,361,648	$7,542,602

Appendix C: Operating Costs

Table C 1: Detailed operating costs for switchgrass fed AD with cotreatment

Production Metrics												
Aspen ID	STRM.98CH4	HEAT.A500.LHV-1						STRM.A100.102	STRM.A100.102	DEN.PRD		
Metric	RD Production Rate (kg/hr)	Total Fuel LHV (MMKcal/hr)	Total Fuel Rate	Total Fuel volume		Fuel Yield from Biomass	Feedstock Rate (DW)	Feed mass flow	Feed H2O flow	Fuel Density		
Value	10725.57	120.50	32.47	136.2250785		49.4	657000.00	90000	5400	0.164		
Units	KG/HR	MMKcal/hr	(MM GGE/yr)	MM Gal/yr		GGE/Dry Tonne	Dry Tonnes/Yr	kg/hr	kg/hr	kg/L		

Operational Variables								
Aspen ID								
Metric	Operating Hours	On Stream Factor			Feedstock Moisture	Feedstock Cost	Glucose price	
Value	7884	90%			6.0%	$77.60	$0.29	
Units	hrs/yr	%			wt %	/dry Tonne	$/lb	

Total Operating Costs										Totals:	$68.79	$211.86
Variable Operating Costs												
PATHWAY	Config	Raw Material	ASPEN ID	kg/hr	lb/hr	Quoted Price (cents / tonne)	Year of Price Quote	2014 Cost ($ / tonne)	2014 Cost ($/lb)	$/hour	MM$/yr (2014)	Cents/GGE Fuel (2014)
Feedstock												
Feedstock												
1	1	Feedstock 1	STRM.A100.102	90000.0	198450.00	7294	2011	$77.89	$0.039	$7,728.81	$60.93	188
										Subtotal:	**$60.93**	**188**
Raw Materials												
1	**A100**	**Feed Handling**										
	0	Electricity import		861.0	kW	by year	2007	$/kWh	$0.0682	$58.72	$0.46	1
1	**A200**	**Anaerobic Digestion**										
	0	Electricity import		470.0	kW	by year	2007	$/kWh	$0.0682	$32.05	$0.25	0.778248574
1	**A300**	**Cotreatment**										
	0	Electricity import		675.0	kW	by year	2007	$/kWh	$0.0682	$46.04	$0.36	1.11769742
1	**A400**	**Upgrading**										
	0	Electricity import		6000.0	kW	by year	2007	$/kWh	$0.0682	$409.20	$3.23	9.935088174
1	**A600**	**Utilities**										
	1	Cooling Tower Chems		2.0	4.41	200000	1999	$3,596.53	$1.798	$7.93	$0.06	0.192543094
	1	Makeup Water	STRM.A100.101	1291666.7	2848125.01	20	2004	$0.31	$0.000	$443.04	$3.49	10.756588
										Subtotal:	**$7.86**	**24**
Co-Products and Credits												
1												
	1	Electricity Export		0.0	kW	by year	2007	$/kWh	$0.0572	$0.00	$0.00	0
										Subtotal:	**$0.00**	**0**
Variable Operating Cost Total											**$68.79**	**212**

Fixed Operating Cost												
		Position	Salary	Year of salary quote	Salary (2014)	# Required	Total Cost (2014)				MM$/yr (2014)	Cents/GGE Fuel (2014)
Labor and Supervision												
		Plant Manager	$147,000	2009	$155,617	1	$155,617					
		Plant Engineer	$70,000	2009	$74,103	2	$148,207					
		Maintenance Supr	$57,000	2009	$60,341	1	$60,341					
		Maintenance Tech	$40,000	2009	$42,345	12	$508,138					
		Lab Manager	$56,000	2009	$59,283	1	$59,283					
		Lab Technician	$40,000	2009	$42,345	2	$84,690					
		Lab Tech-Enzyme	$40,000	2009	$42,345	2	$84,690					
		Shift Supervisor	$48,000	2009	$50,814	4	$203,255					
		Shift Operators	$40,000	2009	$42,345	20	$846,897					
		Shift Oper-Enzyme	$40,000	2009	$42,345	8	$338,759					
		Yard Employees	$28,000	2009	$29,641	4	$118,566					
		Clerks & Secretaries	$36,000	2009	$38,110	3	$114,331					
		Total Salaries					$2,722,772				$2.72	8
		Labor Burden	90.0%	of Total Salaries			$2,450,495				$2.45	8
										Subtotal:	**$5.17**	**16**
Overhead												
		Maintenance	3.0%	of ISBL			$2,682,608				$2.68	8
		Property Insur. & Tax	0.7%	of FCI			$1,532,980				$1.53	5
										Subtotal:	**$4.22**	**13**
Fixed Operating Cost Total											**$9.39**	**29**
Total Operating Costs											**$78.18**	**$240.77**

Appendix D: Discounted Cash Flow Analysis Worksheet

Table D 1: Detailed discounted cash flow analysis worksheet for switchgrass fed AD with cotreatment

Year	Cumulative	-2	-1	0	1	2	3	4
Fixed Capital Investment		$7,007,909	$52,559,318	$28,031,636				
Land		$1,848,000						
Working Capital				$10,949,858				
Loan Payment					$19,582,221	$19,582,221	$19,582,221	$19,582,221
Loan Interest Payment		$840,949	$7,148,067	$10,511,864	$10,511,864	$9,786,235	$9,002,556	$8,156,183
Loan Principal		$10,511,864	$89,350,841	$131,398,296	$122,327,939	$112,531,953	$101,952,288	$90,526,251
Annual Income								
Fuel Sales					$82,114,245	$109,485,660	$109,485,660	$109,485,660
By-Product Credit					$0	$0	$0	$0
Total Annual Sales					$82,114,245	$109,485,660	$109,485,660	$109,485,660
Annual Manufacturing Cost								
Feedstock					$45,700,481	$60,933,975	$60,933,975	$60,933,975
Raw Materials					$6,877,632	$7,860,150	$7,860,150	$7,860,150
Waste Disposal					$0	$0	$0	$0
Fixed Operating Costs					$9,388,856	$9,388,856	$9,388,856	$9,388,856
Periodic Costs (e.g. Catalyst)					$0	$0	$0	$0
Total Product Cost					$61,966,968	$78,182,981	$78,182,981	$78,182,981
Annual Depreciation								
General Plant Writedown					14.29%	24.49%	17.49%	12.49%
Depreciation Charge					$31,294,694	$53,632,404	$38,302,603	$27,352,745
Remaining Value					$187,702,466	$134,070,061	$95,767,458	$68,414,713
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%
Depreciation Charge					$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0
Annual Tax								
Net Revenue					($21,659,281)	($32,115,960)	($16,002,480)	($4,206,249)
Losses Forward						($21,659,281)	($53,775,241)	($69,777,721)
Taxable Income					($21,659,281)	($53,775,241)	($69,777,721)	($73,983,970)
Income Tax					$0	$0	$0	$0
Annual Cash Flow								
Annual Cash Income					$565,056	$11,720,459	$11,720,459	$11,720,459
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830
Annual Present Value					$513,687	$9,686,329	$8,805,754	$8,005,231
Total Capital Investment + Interest		$11,733,198	$65,678,124	$49,493,358				
Cumulative Present Value	**$126,171,254**							
Cumulative Investements + Interest	**$126,171,254**							
Net Present Worth	**$0**							

5	6	7	8	9	10	11	12	13	14	15	16
$19,582,221	$19,582,221	$19,582,221	$19,582,221	$19,582,221	$19,582,221	$0	$0	$0	$0	$0	$0
$7,242,100	$6,254,890	$5,188,704	$4,037,223	$2,793,623	$1,450,535	$0	$0	$0	$0	$0	$0
$78,186,130	$64,858,799	$50,465,282	$34,920,284	$18,131,686	$0	$0	$0	$0	$0	$0	$0
$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660
$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975
$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981
8.93%	8.92%	8.93%	4.46%								
$19,556,446	$19,534,547	$19,556,446	$9,767,273								
$48,858,266	$29,323,720	$9,767,273	$0								
5.71%	5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$4,504,133	$5,513,242	$6,557,529	$17,498,184	$28,509,057	$29,852,145	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679
($73,983,970)	($69,479,837)	($63,966,595)	($57,409,065)	($39,910,882)	($11,401,825)	$0	$0	$0	$0	$0	$0
($69,479,837)	($63,966,595)	($57,409,065)	($39,910,882)	($11,401,825)	$18,450,319	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679
$0	$0	$0	$0	$0	$6,457,612	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938
$11,720,459	$11,720,459	$11,720,459	$11,720,459	$11,720,459	$5,262,847	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742
0.6209	0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176
$7,277,483	$6,615,893	$6,014,448	$5,467,680	$4,970,619	$2,029,055	$7,131,409	$6,483,099	$5,893,726	$5,357,933	$4,870,848	$4,428,044

17	18	19	20	21	22	23	24	25	26	27	28	29	30
													($1,848,000)
													($10,949,858)
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660	$109,485,660
$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975	$60,933,975
$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150	$7,860,150
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856	$9,388,856
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981	$78,182,981
4.46%	4.46%	4.46%	4.46%	2.23%									
$0	$0	$0	$0	$0									
$0	$0	$0	$0	$0									
$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679	$31,302,679
$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938	$10,955,938
$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742	$20,346,742
0.1978	0.1799	0.1635	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
$4,025,494	$3,659,540	$3,326,855	$3,024,413	$2,749,467	$2,499,515	$2,272,287	$2,065,715	$1,877,923	$1,707,203	$1,552,002	$1,410,911	$1,282,647	$1,166,042
													($733,427)

Appendix E: Mill Scale-up Calculations

Source: Personal communication with IKA sales representative and IKA user manual.

Table E 1: Scale and price of different IKA colloid mill equipment

IKA Mill	**Throughput**	**Power**	**Price**
	l/h	*kW*	*USD $*
Labor Pilot 2000/04 (MK Module)	300	1.5	27000
MK 2000/05	2500	4	36000
MK 2000/10	7500	7.5	58000
MK 2000/20	20000	15	100000
MK 2000/30	40000	30	160000
MK 2000/50	80000	75	250000

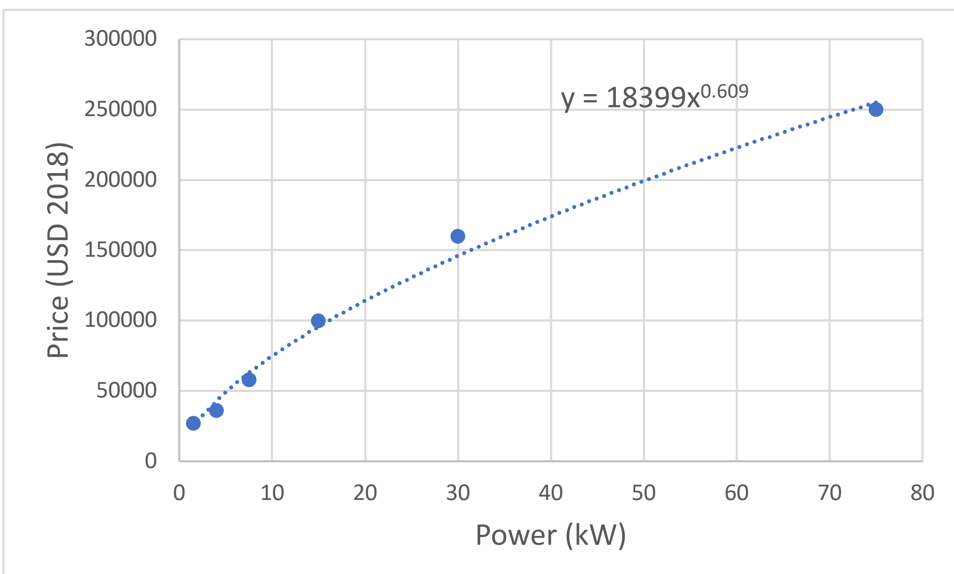

Figure E 1: Power curve fit for IKA mill

Appendix F: Cost Indices

Table F 1: Chemical Engineering Plant Cost Indices

Year	Value
2000	394.1
2001	394.3
2002	395.6
2003	402
2004	444.2
2005	468.2
2006	499.6
2007	525.4
2008	575.4
2009	521.9
2010	550.8
2011	585.7
2012	584.6
2013	567.3
2014	576.1
2015	556.8
2016	541.7
2017	567.5

Source: Chemical Engineering, 2019

Table F 2: Labor Index

Year	Value
2000	17.09
2001	17.57
2002	17.97
2003	18.5
2004	19.17
2005	19.67
2006	19.6
2007	19.55
2008	19.5
2009	20.3
2010	21.07
2011	21.46
2012	21.45
2013	21.4
2014	21.49
2015	21.76
2016	22.71
2017	24.29
2018	25.46

Source: U.S. Department of Labor and Bureau of Labor Statistics, 2019

Table F 3: Electricity Index

Year	Value
2000	4.64
2001	5.05
2002	4.88
2003	5.11
2004	5.25
2005	5.73
2006	6.16
2007	6.39
2008	6.96
2009	6.83
2010	6.77
2011	6.82
2012	6.67
2013	6.89
2014	7.10
2015	6.91
2016	6.76
2017	6.88
2018	6.93

Source: U.S. Energy Information Administration, 2018