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Ming-Hsun Cheng, Zhaoqin Wang, Bruce S. Dien, Patricia J. W. Slininger, Vijay Singh

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Article

Economic Analysis of Cellulosic Ethanol Production from Sugarcane Bagasse Using a Sequential Deacetylation, Hot Water and Disk-Refining Pretreatment

Ming-Hsun Cheng [1], Zhaoqin Wang [1], Bruce S. Dien [2], Patricia J. W. Slininger [2] and Vijay Singh [1,*]

[1] Department of Agricultural and Biological Engineering, University of Illinois at Urbana-Champaign, Urbana, IL 61801, USA; minghsun@illinois.edu (M.-H.C.); wangzhaoqinhan@gmail.com (Z.W.)

[2] Bioenergy Research Unit, National Center for Agricultural Utilization Research, USDA-ARS, Peoria, IL 61604, USA; bruce.dien@ars.usda.gov (B.S.D.); pat.slininger@ars.usda.gov (P.J.W.S.)

* Correspondence: vsingh@illinois.edu; Tel.: +1-217-333-9510

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Abstract: A new process for conversion of sugarcane bagasse to ethanol was analyzed for production costs and energy consumption using experimental results. The process includes a sequential three-stage deacetylation, hot water, and disk-refining pretreatment and a commercial glucose-xylose fermenting *S. cerevisiae* strain. The simultaneous saccharification and co-fermentation (SScF) step used was investigated at two solids loadings: 10% and 16% w/w. Additionally, a sensitivity analysis was conducted for the major operating parameters. The minimum ethanol selling price (MESP) varied between $4.91and $4.52/gal ethanol. The higher SScF solids loading (16%) reduced the total operating, utilities, and production costs by 9.5%, 15.6%, and 5.6%, respectively. Other important factors in determining selling price were costs for fermentation medium and enzymes (e.g., cellulases). Hence, these findings support operating at high solids and producing enzymes onsite as strategies to minimize MESP.

Keywords: sugarcane bagasse; sequential three-stage pretreatment; simultaneous saccharification and co-fermentation (SScF); production cost; minimum ethanol selling price

1. Introduction

Currently, transportation is the largest generator of greenhouse gas (GHG) emissions and energy and consumption by this sector is estimated to increase 60% by 2030, in part due to global population growth [1]. In this regard, bioethanol is considered a promising, environmentally friendly alternative for petroleum-based fuels. The most abundant feedstock for bioethanol production is in the form of lignocellulose (e.g., cellulosic biomass). Globally, the energy supply generated from cellulosic materials is estimated to be 60 EJ, accounting for greater than 10% of the total annual energy supply [2].

Numerous types of cellulosic materials have been investigated as feedstocks for ethanol production, including crop residues (corn stover, sugarcane bagasse, wheat straw), herbaceous biomass (switchgrass, prairie cordgrass, hay), hardwood (poplar, pine) and municipal solid wastes (office paper). Generally, lignocellulose contains 55% to 75% carbohydrates (cellulose and hemicellulose) and 10% to 30% lignin. The complex plant cell wall structure is recalcitrant to extraction of carbohydrates and their depolymerization to monosaccharides, as is necessary for subsequent ethanol fermentation. Therefore, pretreatment is essential for enzymatic hydrolysis [3]. Pretreatments are graded according to multiple targets, including degree of particle size reduction, fermentable sugar (hexose/pentose) recovery, inhibitor formation (e.g., "fermentability"), energy and water consumption, as well as operating cost and environmental footprint.

Dilute-acid pretreatment is one of the most effective methods to reduce biomass particle size and increase cellulose accessibility for cellulases. However, the use of harsh chemicals leads to sugar losses from degradation, formation of fermentation inhibitors, additional chemical usage for post-pretreatment pH adjustment, and corrosion of equipment, which increases operating costs and creates environmental concerns. On the other hand, mild pretreatments, such as liquid hot water (LHW), avoid the formation of excess inhibitors that would otherwise require conditioning hydrolysates prior to fermentation. The major disadvantage of pretreating with LHW is the low yield of sugars following enzymatic hydrolysis. One strategy is to combine LHW with disk refining. LHW followed by disc refining improved sugar yields compared to either pretreatment used alone in the case of herbaceous biomass; as demonstrated for: rice straw [4], corn stover [5] and sugarcane bagasse [6]. However, adding a disc mill significantly inflates capital costs and pretreatment is already the single most expensive unit operation [7]. Therefore, determining the full merits of LHW disc refining requires an economic analysis.

Numerous pretreatments have been researched over the past decades. The pretreatment technology chosen affects operating costs related to utility consumption (steam, electricity), labor and wastewater treatment. Feedstock and chemical costs are similar among various pretreatment options and these make up 60% of the total production costs [8]. An economic comparison of leading pretreatment technologies, using the same process assumptions, found that LHW pretreatment has higher production cost than dilute acid and ammonia fiber expansion (AFEX) due to lower ethanol yields and higher energy inputs [8,9]. However, the pretreatment operating conditions play an important role in determining production cost, especially utility cost. As reported by da Silva (2016) the total energy usage of LHW pretreatment increased from 232.3 to 393.6 MW when pretreatment solids loading (SL) decreased from 20% to 13% [10]. Heat accounts for over 70% of the total energy requirement in the LHW process because high temperatures are required to achieve targeted sugar yields. After feedstock, enzymes (e.g., cellulase) are the next largest material cost item (>20%) [9,11]. Onsite enzyme production is considered as a promising approach to reduce enzyme costs [8], albeit an onsite enzyme production facility inflates fixed capital costs by 13%.

As reported by Kazi et al., (2010), the minimum ethanol selling price (MESP) ranges from $2 to $7/gal depending on pretreatment and fermentation strategies [9]. Basic (mostly AFEX) and LHW (i.e., hydrothermal and liquid hot-water) pretreatments have higher MESP ($3.69 to $5.08/gal) than dilute acid pretreatments ($3.40 to $4.38/gal) [9,12]. However, there are other factors that could affect ethanol production costs, including solids content during hydrolysis and fermentation, fermentation technologies, material and equipment suppliers, and biorefinery site selection [8]. Therefore, under a very conservative scenario, the MESP for dilute-acid pretreatment increased to $7.08/gal [9]. These results were for switchgrass, a report on ethanol production from sugarcane bagasse using dilute acid steam explosion pretreatment estimated the MESP as $1.91 to $2.37/gal [13].

Chen (2015) [14] reported a process that combined dilute alkaline deacetylation and disk refining (DDR) that achieved high sugars and ethanol titers. Economic analysis of the DDR process predicted a minimum ethanol selling price (MESP) of 2.24 to $2.54/gal ethanol, for the case of enzymatic hydrolysis at 15% solids. The majority of cellulosic ethanol studies operated at low solids (5% to 10% w/w) saccharification and fermentation that favor high yields. However, realistically for distillation, ethanol concentrations should be greater than 4% [15], which requires high solids (≥15% w/w) loading hydrolysis and fermentation [16]. The objective of this study was to evaluate the economics feasibility of a new cellulosic ethanol process that implements deacetylation followed by combined LHW-disc refining of sugarcane bagasse.

2. Materials and Methods

2.1. Process Description

The cellulosic ethanol process using sugarcane bagasse was designed to process 2000 metric tonnes/day. The detailed process model was modified from NREL (National Renewable Energy Laboratory)/TP (technical report)-6A2-46588 [17] and ethanol production from sugarcane bagasse with sequential pretreatment and simultaneous saccharification and co-fermentation reported by Wang et al., (2019) [18]. The process includes biomass handling, deacetylation, hot water pretreatment, disk milling, onsite enzyme production, saccharification and co-fermentation, ethanol distillation and steam generation (Figure 1).

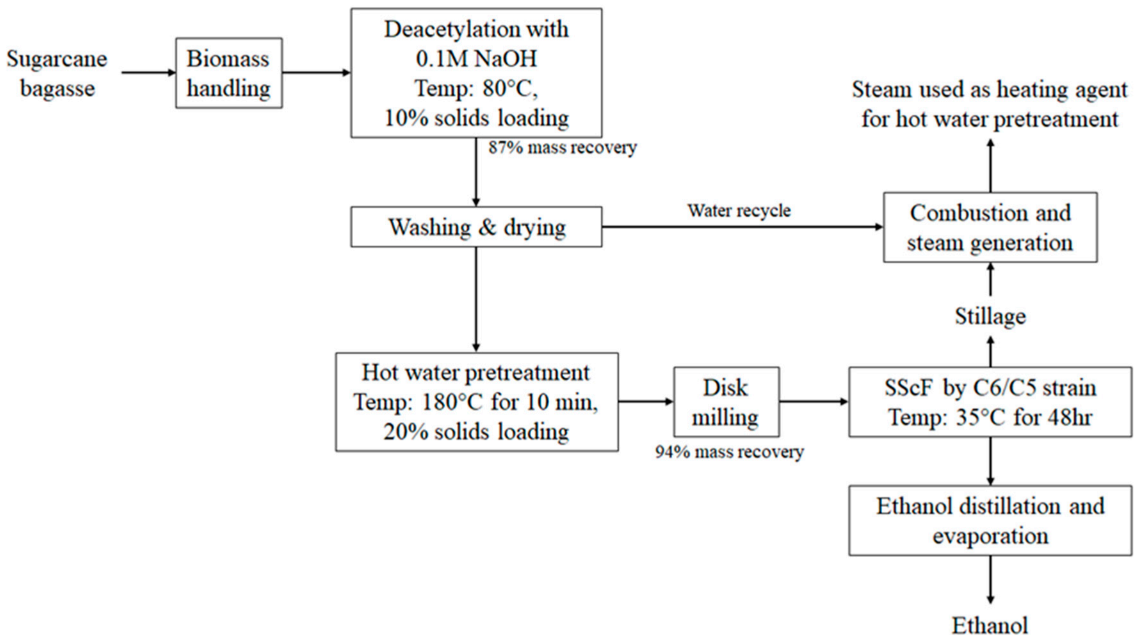

Figure 1. Overall workflow for production of cellulosic ethanol production using a sequential deacetylation, hot water and disk-milling pretreatment process.

2.1.1. Sugarcane Bagasse

The dried sugarcane bagasse contains (%w/w, oven dry basis): 37% glucan, 20% xylan, 20% lignin, 8% extractives, 4% ash, and 2% acetyl group [18]. The amounts of glucan and xylan were used to calculate the ethanol conversion efficiency based on the theoretical ethanol yields.

2.1.2. Biomass Handling and Deacetylation

Harvested sugarcane bagasse was transported to the mill, where the biomass was shredded to 2.0 mm to improve the sugar release and ethanol yields in subsequent processes. Impurities were removed from the shredded bagasse by using a magnetic separator.

The shredded and cleaned biomass was deacetylated by soaking in 0.1M NaOH at 10% solids loading at 80 °C for 3 h with continuous agitation at 70 rpm. The chemical compositions of sugarcane bagasse before and after deacetylation process was reported by Wang et al., (2019) [18] and listed in Table 1. In the deacetylation process, 87.63% of mass was recovered, about 65.75% acetyl group, 0.58% of AIL (acid insoluble lignin) and 0.3% ASL (acid soluble lignin) were removed by alkaline treatment. Additionally, xylan content was reduced from 20.40% to 15.88% due to removal of acetyl group. The deacetylated biomass was washed by water at a 1:1 solids:liquid ratio. In the washing process, about 0.63% of ash was removed. The waste-water was recycled for steam generation in the utility generation section. The washed biomass was fed into the hot water pretreatment reactor.

Table 1. Composition (% w/w, dry basis) of raw and dacetylated sugarcane bagasse.

	Mass Recovery	Extractive	Glucan	Xylan	AIL [a]	ASL [b]	Ash	Acetyl Group
Raw	NA	7.82 ± 0.88	37.31 ± 1.72	20.40 ± 2.61	18.18 ± 0.48	2.22 ± 0.04	3.65 ± 0.47	2.16 ± 0.18
Deacetylated	87.63 ± 0.48	NA	34.40 ± 0.88	15.88 ± 1.32	17.60 ± 1.91	1.92 ± 0.08	3.02 ± 0.36	0.74 ± 0.02

[a] AIL: acid insoluble lignin (Klason lignin). [b] ASL: acid soluble lignin.

2.1.3. Hot Water Pretreatment and Disk Refining

The sugarcane bagasse was pretreated using hot water pretreatment with 20% solids loading at 180 °C for 10 min [6]. The reactor was heated by high pressure steam. Because hydrothermal pretreatment was at 180 °C, there were no furans generated from glucose and xylose degradation [18]. After the hot water pretreatment, the unwashed biomass was disk milled. The energy consumption for disk milling was set at 315 kwh/ODMT (oven dried metric tonne) based on Chen et al.'s study (2015) [14]. The mass loss of bagasse from disk milling was set to 6% as per laboratory results [18].

2.1.4. Simultaneous Saccharification and Co-Fermentation (SScF)

After the hot water pretreatment and disk milling, the pretreated biomass was cooled and fed into the fermenter. The C5/C6 yeast M11205 seed (Lallemand Inc., Milwaukee, WI, USA) was cultured in a medium containing 40% glucose at 32 °C. Additionally, cellulase used in the process was assumed to be produced on-site using the seed culturing facility.

The SScF was carried out with 10% w/w substrates loading (pH 5, 32 °C for 72 h), in which cellulase Cellic® Ctec2 (protein content: 76.3 mg/mL by bovine serum albumin (BSA) analysis) (Novozymes North America, Inc., Franklinton, NC, USA) at 0.215 mL/g dry substrate basis (in fermentation broth) were applied to ferment hexoses and pentoses into ethanol. The conversion factor of glucan to glucose and xylan to xylose were 85% and 80% (based on raw bagasse chemical compositions), respectively, and the fermentation efficiency from sugars to ethanol was 94.3% [18].

2.1.5. Ethanol Distillation and Evaporation

Ethanol is recovered and purified subsequently from the fermentation beer using a combination of distillation columns and molecular sieves. In the distillation process, ethanol is first separated in a distillation column as overheated ethanol-enriched vapors, which contain equal amounts of ethanol and water. The ethanol purity is further enriched in the rectification and stripper columns. However, an azeotrope mixture of water and ethanol forms in the rectification and stripper columns, and ethanol cannot be separated using distillation. Molecular sieves are used to separated ethanol from the azeotrope mixture to a final ethanol purity of >99% [19,20].

2.1.6. Steam Production

Solid residues after fermentation, known as stillage, consist mainly of lignin and unfermented structural carbohydrates, which are burned and the heat used to generate process steam with the water recycled from the washing step after deacetylation. In this section, a combustor and a boiler are the major pieces of equipment. The stillage fed into the combustor contains 50% moisture. The process design and technical inputs used in this section are based on the cellulosic ethanol model developed by the National Renewable Energy Laboratory (NREL) [9,17]. The heating value of the stillage was calculated according to the element composition using the software model-embedded combustor module. The efficiency of boiler for steam generation was set at 80% [20]. The steam is extracted at 1.48 MPa and 268 °C, which is used in hot water pretreatment process. Because the total steam generated from stillage in this section cannot meet the steam requirement for the whole process, especially for the hot water pretreatment step, stream was not used to generate electricity unlike the NREL model.

2.1.7. Scenarios

In this study, solids loadings of 10% and 16% in the fermentation were investigated according to the experimental data reported by Wang et al., (2019) [18]. For these two scenarios, the plant capacity was kept the same (2000 metric tonnes/day of sugarcane bagasse).

2.2. Economic Analysis

Economics of the three stage deacetylation, hot water, and disk milling pretreatment for ethanol production from sugarcane bagasse was estimated using SuperPro Designer (Version 9.5, Intelligen Inc., Scotch Plains, NJ, USA, 2017) process modeling software (Supplementary Figure S1). The major economic parameters are listed in Table 2. The plant was assumed to be located in Illinois. For each unit operation, relevant processing equipment were selected from the SuperPro Designer database. However, some processing units were modelled using component splitters because of its unavailability in the software library. The main non-design parameters used for the economic profitability analysis were selected from the NREL/TP-6A2-46588 [9,17] report on ethanol production from lignocellulosic biomass and Huang et al. (2016) [19].

Table 2. Major economic parameters for analysis of ethanol production from sugarcane bagasse.

Parameter	Value
Project lifetime	20 years
Salvage value of equipment	0
Construction and startup time	2 years
1st year TCI [a] allocation	40%
2nd year TCI allocation	60%
Depreciation life	10 years with straight-line method
Income tax	35%
Interest	10%

[a] TCI: total capital investment.

2.2.1. Fixed Capital Investment

Equipment costs were used to estimate total fixed capital cost according to literature [17,19]. Where equipment capacity was re-scaled from prior studies, an exponential scaling expression (n = 0.4 to 0.8) was used with the power term set according to literature (Equation (1)). The equipment costs used in this analysis are listed in Table 3 and are adjusted to 2017 US dollars using a standard inflation index.

$$New\ cost\ =\ (Base\ cost)\left(\frac{New\ scale}{Base\ scale}\right)^n \tag{1}$$

The total fixed capital cost was calculated by multiplying the machine purchase cost by the Lang factor, which is the estimated ratio of the fixed capital cost (e.g., installation) to the machine purchase fee. Capital costs included plant direct costs (installation, process piping, instrumentation etc.), plant indirect costs (engineering and construction fees), and contractor's and contingency fees. The Lang factor was set to 3, which is the standard value for a biorefinery facility [17,19,21]. Working capital and startup cost were set to 20% of the total fixed cost. Total fixed capital investment was the sum of the fixed capital, working capital and startup costs.

Table 3. Main equipment cost of the ethanol production form sugarcane bagasse (2017 price).

Process Section	Equipment	Base Scale	Base Cost	New Scale	n *	Adjusted Cost
Biomass handling	Shredder	300 MT/h	1,932,000	100 MT/h	0.6	1,001,000
	Conveyor	300 MT/h	333,900	100 MT/h	0.6	334,000
	Magnetic separator	720 MT/h	30,000	100 MT/h	0.6	9,000
Deacetylation	Blending tank	250,000 gal	4,032,000	925,000 gal	0.6	9,597,000
	Centrifuge +	n/a	n/a	90,000 hL/h	0.49	253,000
	Washing tank +	n/a	n/a	420 MT/h	0.6	1,833,000
Hot water pretreatment	Reactor #	83.33 MT/h	22,585,680	167 MT/h	0.6	34,028,000
	Cooling system	400 m^2	194,250	3000 m^2	0.8	974,000
Disk milling	Disk mill	420 MT/h	315,000	370 MT/h	0.6	292,000
Saccharification and fermentation	Seed fermenter #	300 m^3	456,570	240,000 gal	1	1,381,000
	Centrifuge	1200 MT/h	569,100	1200 MT/h	0.49	569,100
	Blending tank #	200 m^3	131,100	1500 m^3	0.5	359,000
	Fermenter	1000 m^3	657,300	1900 m^3	0.6	996,100
	Gas absorber	4 m	1,668,000	4m	0.7	1,668,000
Ethanol distillation	Disk-stack centrifuge	1200 MT/h	569,100	800 MT/h	0.49	467,000
	Distillation column	4 m	505,000	4 m	0.6	505,000
	Rectifier column	2 m	110,250	4 m	0.6	167,000
	Stripping column	2 m	198,000	2 m	0.6	198,000
	Molecular sieve	25 MT/h	2,222,850	42 MT/h	0.6	3,034,200
	Heat exchanger	200 m^2	52,500	750 m^2	0.6	116,000
	Product tank	14,158 m^3	408,450	5820 m^3	0.7	219,000
	Ethanol pump	720 MT/h	5250	720 MT/h	0.8	5,250
Utility (steam generation)	Evaporator	900 m^2	2,186,000	900 m^2	0.7	2,186,000
	Steam boiler	400 MT/h	33,297,000	100 MT/h	0.6	14,496,000

The data of base scale are collected from Huang et al.'s study (2016) [19], Humbird et al.'s study (2011) # [17], and SuperPro Designer database +. * n: exponential scaling factor.

2.2.2. Operating Cost

Production cost included variable and fixed operating costs. Materials and utilities account for most of the variable operating costs. Fixed operating costs include labor and facility-related maintenance fees, which are incurred in full regardless of the production rate. For labor cost, eight operators and two laboratory technicians are assigned to pretreatment, seed culture, ethanol fermentation and distillation operations. All operating costs are listed in Table 4.

Table 4. Operating cost inputs.

Costs		Cost	Citation
Feedstock	Sugarcane	$0.05/kg	Kazi et al., 2010 [9]
Chemicals	NaOH	$0.15/kg	SuperPro Designer *
	Citrate buffer (1M)	$0.01/L	
	Glucose	$0.56/kg	
	$(NH_4)_2SO_4$	$0.96/kg	
Utility	Electricity	$0.07/kwh	US EIA [22]
	Steam	$2.6/MT	US DOE [23]
	High pressure steam	$20/MT	US DOE [23]
	Water	$0.001/L	Cheng et al., (2017) [24]
Labor	Operator	$15.12/h	US BLS [25]
	Fermentation operator	$22.79/h	
	Laboratory technician	$22.87/h	
Facility related	Maintenance	7% of facility purchase price	Cheng et al., (2017) [24]

* SuperPro Designer: date input were collected from SuperPro Designer database. US EIA: US Energy Information Administration; US DOE: US Department of Energy; US BLS: US Bureau of Labor Statistics.

2.3. Sensitivity Analysis

Economic models are valuable in predicting the commercial feasibilities of new technologies. They can also help to select among different designs and determine sensitivities to design changes, changes in material costs, or when targets are not fully achieved.

The efficiency of pretreatment (energy usage, fermentable sugar yields) and solids loading used in fermentation determine the final ethanol yield. Sensitivity analysis was performed by varying energy consumption used in disk milling, hydrolysis efficiency and solids loading to the fermentation tank to investigate how these affect ethanol yield and profitability. Feedstock costs may be difficult to predict and will vary depending upon location. Therefore, feedstock cost (±40% based on the base scenario) is also included in the sensitivity analysis.

There are no reports in literature regarding enzyme production costs when manufactured onsite. Therefore, a purchase cost estimation was substituted. According to the study from Gubicza et al., (2016), the purchased enzymes for cellulosic ethanol production is $1/kg enzyme [13].

3. Results and Discussion

3.1. Fermentation Efficiecncy and Ethanol Yields

An economic analysis of sugarcane bagasse to ethanol was conducted using prior research results using the three-stage pretreatment and SScF process. The SScF was operated at 10% and 16% solids and all other operating parameters were kept constant. Fermentation results from these two processes are summarized in Table 5. Low solids loading (10%) achieved the highest ethanol yield (0.33 g/g biomass) and conversion efficiency (94%). At 16% solids, the ethanol conversion efficiency was reduced to 91%. However, high solids loading is required for fermentation to achieve the minimum ethanol concentration needed for efficient distillation. Therefore, two ethanol efficiencies and responding solids loading (10% and 16%) were used to compare the effect of solids loading on the minimum ethanol selling price (MESP).

Table 5. Fermentation efficiency and ethanol yields.

	Exp.	BS	HS
SScF time (h)	96	72	72
Solid loadings (%)	10	10	16
Ethanol conversion (%)	94.33 ± 2.84	94	91
Ethanol yields (g/g of pretreated biomass)	0.343 ± 0.009	0.33	0.30

Exp: experimental data (Wang et al., 2019) [18]; BS: base scenario; HS: higher solids loading.

3.2. Fixed Capital Investment of Process Plant

Fixed capital cost for a plant with an annual capacity of 54.2 million gal was calculated to be $398.94 million, where the majority of the investment comes from unit operations for (listed in order): SScF (42.99%), LHW pretreatment (28.17%), steam generation (14.84%), deacetyaltion (9.64%), distillation (3.37%), biomass handling (0.78%), and disk milling (0.21%) (Figure 2).

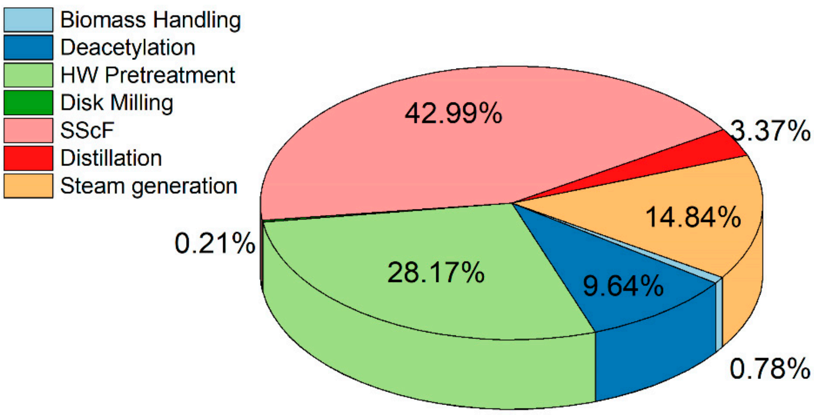

Figure 2. Breakdown of the total fixed capital investment.

Previous studies reported fixed capital costs for production of cellulosic ethanol are from $185 to $253 million [9,17,26], which is lower than that for the three-stage pretreatment process plant. This increase in capital cost is expected because additional processing units were included in this model. The deacetylation and disk milling process increased total capital costs by 9.64% ($35.05 million) and 0.21% ($0.76 million), respectively. Chen et al. (2015) reported the additional high temperature pretreatment reactor increased the total capital investment by $30.45 million (2017 dollars) [14]. In this study, the cost of the high temperature pretreatment reactor was $102 million.

The SScF included seed culturing for enzyme production and ethanol fermentation. In ethanol fermentation, 1000 m^3 fermenters are commonly used. Same size fermenters were used in this study, and there were 36 fermenters required. Additionally, the hot water pretreatment does not require extra costs for anti-corrosion, neutralization, and waste acid water compared to typical acid pretreatment. Therefore, SScF is the major contributor for the fixed capital investment. For distillation, the 3.37% of the total fixed capital investment is comparable to 3–4% of total fixed capital investment for cellulosic ethanol plant reported by Gregg et al., (1998) [27] and 1% reported by Humbird et al., (2011) [17].

The high-solids (HS) scenario, the biomass processing capacity is same as the base scenario (BS) and the major difference is the solids fermented and less water loadings in the bioreactor. Hence, the total fermentation volume of HS is smaller than BS resulting in $41.55 million lower in the fermentation section.

3.3. Operating Costs

Annual operating cost was estimated to be $243.89 million and $220.75 million for BS and HS, respectively; these costs are broken down in Figure 3. Materials and utilities are the largest operating costs, accounting for 80% of the total.

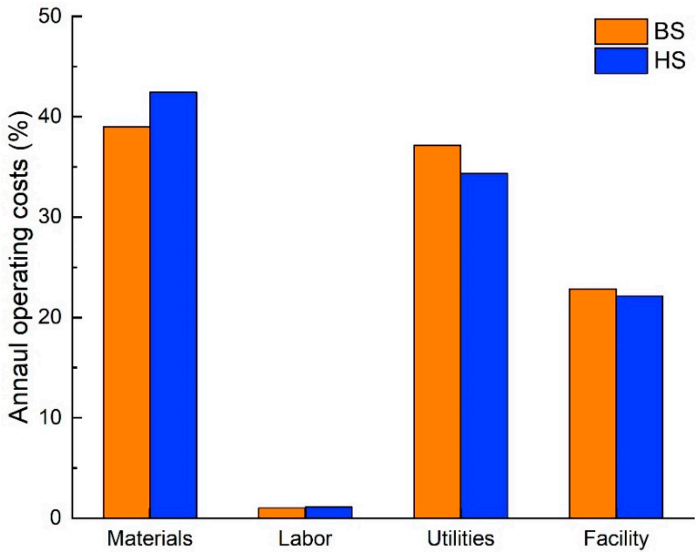

Figure 3. Breakdown of operating costs. BS: base scenario; HS: higher solids loading.

Among the material costs, feedstock bagasse and the seed culturing/fermentation medium account for 35% and 45% of material costs, respectively, for both scenarios. Feedstock is typically the largest cost [8]. However, this process used an engineered glucose and xylose co-fermenting *Saccharomyces* yeast strain and refined/expensive medium was used to propagate the seed culture. For the HS scenario, higher material costs were associated with preparing the inoculum because fermenting a higher concentration of solids requires lager amounts of enzyme and yeast than the BS.

The utility costs were $90.63 and $75.75 million contributing 37.16% and 34.32% of operating cost for BS and HS, respectively, which included normal pressure steam (153 °C, 4.43 bar), cooling water, electricity and high pressure (242 °C, 13.17 bar) steam. Utility costs are itemized in Table 6. Running the LHW pretreatment cost the most (in terms of utilities) followed by distillation and disk milling. The main utility cost associated with LHW pretreatment is high pressure steam, which is the heating agent for the pretreatment reactor. Additionally, after pretreatment, the pretreated biomass needs to be cooled to 100 °C before entering the disk mill. Hence, combined heating and cooling make LHW pretreatment a utility intensive process. Following fermentation, distillation and evaporation require large amounts of steam and cooling water for ethanol recovery.

Table 6. Itemized list of the utility costs of BS and HS scenarios (1000$/year).

	Total Utility		Electricity		Steam		High Pressure Steam		Cooling Water	
Process Section	**BS**	**HS**	**BS**	**HS**	**BS**	**HS**	**BS**	**HS**	**BS**	**HS**
Biomass handling	397	397	397	397	n/a	n/a	n/a	n/a	n/a	n/a
Deacetylation	1091	1091	375	375	716	716	n/a	n/a	n/a	n/a
Hot water pretreatment	32,967	32,967	2	2	n/a	n/a	32,965	32,965	n/a	n/a
Disk milling	8766	8766	8766	8766	n/a	n/a	n/a	n/a	n/a	n/a
SScF	11,289	11,336	2375	2533	n/a	n/a	n/a	n/a	8914	8802
Distillation	33,199	19,146	1911	1135	3871	2341	n/a	n/a	27,418	15,670
Steam generation	2981	2839	276	268	2706	2571	n/a	n/a	n/a	n/a
Total	90,692	76,541	14,103	13,476	7293	5627	32,965	32,965	36,331	24,473

BS: base scenario; HS: higher solids loading.

Electricity is used for grinding, milling, mixing, pumping, and agitation. From the model results, the disk refining consumed over 62% of total electricity followed by saccharification and cofermentation and distillation operations. Disk refining applies shear force (mechanical approach) to defibrillate the biomass at the microscope level; therefore, it consumes large amounts of energy. For SScF and distillation, the electricity is consumed by bioreactor agitators and by pumps that transfer biomass and products between unit operations.

Distillation was the major contributor to cooling water cost, accounting for 75.47% and 64.% in BS and HS, respectively. In this model, pretreated biomass was cooled to 100 °C before disk refining by using a heat exchanger, which would save cooling water in the pretreatment section. The distillation process consisted of beer distillation, ethanol rectification and stripping. Cooling water was required to recover the purified ethanol, which leads to the higher cost. However, the multiple stage pretreatment process accounted for over 75% the total utility cost, arising primarily from steam heating.

Comparing BS scenario and HS scenario, HS has 15.6% ($14 million/year) lower utility costs than BS. Due to the same plant and processing capacity, both scenarios share the same utility costs from biomass handling, deacetylation, hot water pretreatment, and the disk-milling process. In seed culturing and SScF process, HS scenario required more electricity for production of enzymes and yeast. However, the higher ethanol concentration lowers the utility costs associated with ethanol distillation. Therefore, operating at higher solids (HS) lowers the utility costs by 40% (electricity and steam) for distillation, which is the reason that operating at high solids is favored by industry.

3.4. Production Cost and the Minimum Ethanol Selling Price (MESP)

For BS, a yield of 205.06 million L (54.17 million gal) is produced yearly from 2000 MT of dry sugarcane bagasse using this multiple-stage pretreatment process. The 10% solid loading was the baseline for estimating the production cost and the MESP of the ethanol from sugarcane bagasse. The gross production cost was calculated by dividing annual operating costs by ethanol production, which for BS was $4.48/gal ethanol.

Typically, excess heat (generated from combustion of lignin enriched residues) and process waste-water are expected to be used for cogeneration of electricity at cellulosic ethanol plants. In this three-stage pretreatment scenario, the water used in the deacetylation and washing steps was reused for steam generation. However, the steam generated from the steam boiler (650,214 MT high pressure steam/year) cannot satisfy the total steam consumed (1,648,250 MT high pressure steam/year) in the HW pretreatment (180 °C for 10 min), resulting in no surplus energy left for electrical cogeneration. The steam generation and recycled water save operating costs, and net production cost is calculated by subtracting savings from total operation costs. Therefore, the net production cost was $4.16/gal ethanol and the MESP was estimated as $4.90/gal ethanol.

Cost and MESP associated with the HS scenario (16% solids in fermentation) are listed in Table 7. Though the ethanol yield from 16% solid loading in fermentation was lower than the 10% solid loading, total operating costs were decreased by over $23 million/year. The largest cost savings was for utilities, which were reduced by 15% (from $90.7 million/year to $76.5 million/year). This utility cost reduction was mainly from the distillation process due to the high ethanol titer of the beer. Thus, the net production cost of ethanol from 16% solid loading in the fermentation was $3.83/gal ethanol and its MESP was $4.23/gal ethanol. MESP was determined when net profit value equals to zero at the internal rate of 10%. Regardless of solids loading, the payback time is 8.91 years.

Table 7. Production cost and minimum ethanol selling price (MESP) of ethanol from sugarcane bagasse.

	Ethanol Yields	**Production Cost**	**Net Production Cost**	**MESP**	**Payback Time**
Scenario	**Million gal/year**	**$/gal**	**$/gal**	**$/gal**	**year**
BS	54.17	4.48	4.16	4.90	8.91
HS	52.22	4.23	3.83	4.52	8.91

BS: base scenario; HS: higher solids loading.

The MESP of cellulosic ethanol was estimated from $2.38 to $5.15/gal ethanol in 2017 dollars from various feedstocks and pretreatment technologies [9,28,29]. A similar analysis for a multiple-step pretreatment approach was investigated by Chen et al., (2015) [14]. Deacetylation of the same treatment condition combined with the disk-refining process yield from 77 to 89 gal ethanol/MT corn stover, oven dry basis, and the MESP was from $2.24 to $2.54/gal ethanol. The costs from LHW pretreatment and the distillation process resulted in a higher MESP in our study. Although the ethanol yield was increased by 5% compared to the Chen et al. study (2015) [14], the steam and cooling costs were higher because of the pretreatment and distillation processes. Hence, utilities is an important area to target to reduce costs.

3.5. Energy Efficiency

Steam and electricity are main energy used to convert biomass to ethanol. The net energy input (NEI) indicates the energy input required in the process to produce ethanol, and it is defined in Equation (2), where E_p, E_s, E_{hs} and E_e indicate the energy content of ethanol product, steam, high pressure steam and electricity, respectively. Additionally, the energy ratio (Equation (3)) represents the energy efficiency of converting biomass to ethanol. It is measured by dividing NEI by the energy content of sugarcane bagasse (15.6 MJ/kg) [30]. An energy ratio greater than one means more energy inputs, higher than biomass energy content, are required for the process to produce ethanol.

$$\text{NEI} = (E_s + E_{hs} + E_e) - E_p \tag{2}$$

$$\text{Energy Ratio} = \frac{\text{NEI}}{\text{Energy content of sugarcane bagasse}} \tag{3}$$

Table 8 lists the net energy and energy ratio derived from BS and HS scenarios. BS has higher total energy input than HS in steam and electricity, which were from distillation process. Though BS has higher ethanol conversion rate, the lower ethanol titer resulted in higher energy consumption in the ethanol recovery. For net energy and energy ratio, HS is about 17% lower than BS. However, the energy ratios are 0.95 and 0.78 for BS and HS, respectively, which are higher than 0.61 reported by NREL of ethanol production from corn stover using dilute acid pretreatment [17]. From the results, the high pressure steam, used in LHW pretreatment, is the major contributor to the total energy consumption. Therefore, improving energy efficiency is the main challenge for applying this multistep pretreatment in industrial applications.

Table 8. Net energy and energy ratio for BS and HS scenarios.

	Energy Flow (MJ/h)	
	BS	**HS**
Ethanol	524,300	504,975
Electricity	93,173.04	83,801.92
High-pressure steam	688,851.96	688,851.96
Steam	973,905.56	751,534.38
Net energy input (NEI)	1,231,630.55	1,019,213.25
Sugarcane bagasse	1,300,002.6	1,300,002.6
Energy ratio	0.95	0.78

BS: base scenario; HS: higher solids loading.

3.6. Sensitivity Analysis

Sensitivity analyses were performed based on BS to identify the variables that would most affect bioethanol production costs (Figure 4). Among all the factors analyzed, enzyme costs had the largest effect. The net cost of producing ethanol increased to \$6.36/gal (53% increase in ethanol unit production cost and 48% in total operating costs) if the cellulase was purchased from the market for \$1/kg. However, the enzyme price (\$1/kg enzyme) was also an estimated value; therefore, the 53% increase in the ethanol unit production cost is an approximation, which is 18% higher and 10% lower than the base and pessimistic scenarios reported by Kazi et al., (2010) [9], respectively. From this result, minimizing enzyme costs is critical for cellulosic ethanol plants, which favors on-site enzyme production.

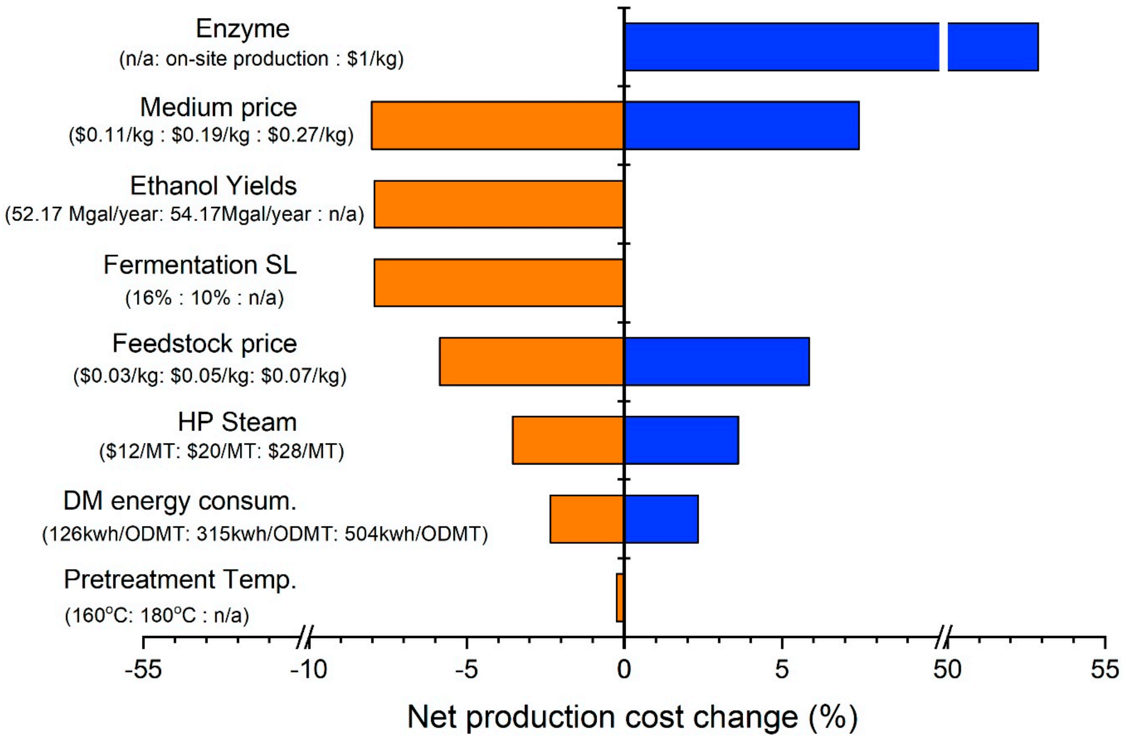

Figure 4. Sensitivity of the ethanol production cost to different parameters.

Product cost was most sensitive to the costs for fermentation solids loading, medium (to grow the yeast inoculum) and feedstock. Production costs varied −8.02% to +7.45% when the unit cost of medium ranged from \$0.11 to \$0.27/kg (\$0.19/kg in BS). However, it was reported that there was alternative yeast medium with lower cost that had a similar performance on poplar SScF [31]. Increasing fermentation solids loading had negative effect on conversion efficiency but increased efficiencies for downstream distillation and evaporation processes. The ethanol yield decreases about 2 million gal/year when increasing solids loading from 10% (54.17 million gal/year) to 16% (52.22 million gal/year). However, higher ethanol titer from 16% solids loading reduced about 42% utility cost in the distillation process which decreased net ethanol production cost from $4.16/gal to $3.83/gal and lowered the MESP from $4.90/gal to $4.52/gal. This indicated that high solids saccharification and fermentation are economically preferred for industrial scale production. We also varied the cost of sugarcane bagasse. The cost of bagasse for the base case is $0.05/kg and varied between $0.03/kg ($30/MT) and $0.07/kg ($70/MT). These changes resulted in cellulosic ethanol net production costs of $3.92 to 4.40/gal.

In addition, the price of high-pressure (HP) steam played an essential role in determining ethanol production costs because of the large amount consumed by this pretreatment process. However, when the unit price of HP stream varied from $12 to 28/MT, the ethanol net production cost changed ±3.6%. Also, energy consumption for disk refining was adjusted ±60% [14]. The ethanol net production cost changed by ±2.33%. Pretreatment temperature was also varied but it had little effect on ethanol production cost. Based on the results, utility cost is higher for this three-stage pretreatment process. For lowering the utility costs, higher solids loading (~50%) in the hot water pretreatment process could be an alternative to improve the energy efficiency. Cheng et al.'s (2019) [32] reported that high solids loading (50%) pretreatment at 180 to 190 °C resulted in increasing biomass surface area by 65% which facilitates the following disk-milling process and increases the sugar yields. Hence, the higher solids loading in hot water pretreatment is expected to lower the total operating cost and MESP.

4. Conclusions

The total fixed capital investment for this three-stage pretreatment ethanol production process was $399 million with LHW pretreatment, which accounted for 29% of that cost following the SScF fermentation unit (43%) followed by the utility equipment (15%). The disk-refining process consumed 7% to 10% of the electricity but improved the sugar yield by 20% and increased ethanol production by 25 L/MT dried biomass. The MESP was 4.52/gal at a solids loading of 16% (HS) and $4.90/gal at a 10% solids loading SScF (BS). Major operating costs were the large amount of steam consumed by the LHW reactor. Other significant costs included the seed-culturing medium, feedstock, and process steam. Therefore, the improvement of heating energy (steam) recovery and reduction in enzyme cost are critical and required for industrial operations to reduce the ethanol production cost and MESP. For improving the energy efficiency in pretreatment, higher solids loading (~50%) in the hot water pretreatment could be an alternative. The economic evaluation of the high solids loading pretreatment process will be investigated and validated in the pilot-scale operations.

Supplementary Materials: The following are available online at http://www.mdpi.com/2227-9717/7/10/642/s1: Figure S1: Economic model of ethanol production from sugarcane bagasse using three-stage pretreatment and SScF.

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