Techno-economic analysis of ethanol production from sugarcane bagasse using a Liquefaction plus Simultaneous Saccharification and co-Fermentation process

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Krisztina Gubicza [a], Ismael U. Nieves [b], William J. Sagues [b], Zsolt Barta [a], K.T. Shanmugam [b,c], Lonnie O. Ingram [b,c,*]

[a] Department of Applied Biotechnology and Food Science, Faculty of Chemical Technology and Biotechnology, Budapest University of Technology and Economics, H-1111 Budapest, Muegyetem rkp. 3, Hungary
[b] Stan Mayfield Biorefinery, University of Florida, Perry, FL 32348, United States
[c] Microbiology & Cell Science, University of Florida, Gainesville, FL 32611-0700, United States

HIGHLIGHTS

- A simplified (L + SScF) process with phosphoric acid was modelled using Aspen.
- A $10 mil savings in capital cost was achieved by using phosphoric acid.
- Capital investment ranged from $169 to $197 mil for various scenarios.
- Ethanol production cost varied between 50 and 63 cents per liter.
- Overall ethanol yield had the biggest impact on production cost.

GRAPHICAL ABSTRACT
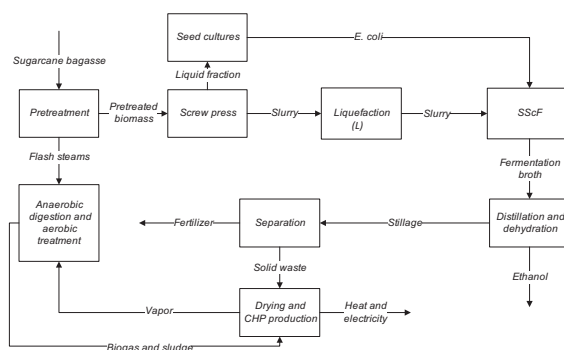
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ABSTRACT

A techno-economic analysis was conducted for a simplified lignocellulosic ethanol production process developed and proven by the University of Florida at laboratory, pilot, and demonstration scales. Data obtained from all three scales of development were used with Aspen Plus to create models for an experimentally-proven base-case and 5 hypothetical scenarios. The model input parameters that differed among the hypothetical scenarios were fermentation time, enzyme loading, enzymatic conversion, solids loading, and overall process yield. The minimum ethanol selling price (MESP) varied between 50.38 and 62.72 US cents/L. The feedstock and the capital cost were the main contributors to the production cost, comprising between 23–28% and 40–49% of the MESP, respectively. A sensitivity analysis showed that overall ethanol yield had the greatest effect on the MESP. These findings suggest that future efforts to increase the economic feasibility of a cellulosic ethanol process should focus on optimization for highest ethanol yield.

1. Introduction

Due to the volatile nature of oil prices and environmental concerns, a great deal of attention has been placed on renewable lignocellulose-based fuels and chemicals to replace current

* Corresponding author at: Department of Microbiology and Cell Science, Bldg 981, University of Florida, Gainesville, FL 32611-0700, United States. Tel.: +1 352 392 8176; fax: +1 352 846 0969.
E-mail address: Ingram@UFL.edu (L.O. Ingram).

petroleum-based products. Initial economic analyses performed on cellulosic fuel ethanol production cited conversion economics as the main issue to be addressed (Lynd et al., 1991), while other techno-economic models focused on optimization of operational costs (Nguyen and Saddler, 1991; von Sivers and Zacchi, 1995; Wyman, 1994). However, significant progress has been made since these earlier studies, and recent techno-economic analyses provide a more favorable view for lignocellulosic ethanol production (Chovau et al., 2013).

The National Renewable Energy Laboratory (NREL) published in 1999 a detailed analysis for lignocellulosic ethanol production and reported an ethanol production cost of 0.38 US$/L (1.44 US$/gal) (Wooley et al., 1999). A second report by NREL in 2002 with revised figures for equipment and installation costs, projected the required advances needed in key research areas with the aim to reach a MESP of 0.28 US$/L (1.07 US$/gal) in 2010 (Aden et al., 2002). Subsequent techno-economic analyses have made use of some of the parameters from the NREL report on the operation of an *n*th plant. Nevertheless, the reported MESP values have varied considerably from one study to the next based on the assumptions and process configurations (Aden and Foust, 2009; Chovau et al., 2013; Eggeman and Elander, 2005; Foust et al., 2009; Galbe et al., 2007; Hamelinck et al., 2005; Han et al., 2015; Kumar and Murthy, 2011; Macrelli et al., 2012; Sassner et al.). These differences have made it difficult to compare these studies (Chovau et al., 2013; Galbe et al., 2007; Sassner et al., 2008). The NREL report (Wright et al., 2010) was further revised with more representative values in 2011 and resulted in a MESP of 0.57 US$/L (2.15 US$/gal) (Humbird et al., 2011).

Some of the significant contributors to the MESP of lignocellulosic ethanol include the cost of the feedstock, the ethanol yield, and the cost of cellulase enzymes (Chovau et al., 2013). However, the main contributor to the MESP in almost all cases seems to be the capital cost (Galbe et al., 2007). From various studies, it is clear that one way to lower the MESP is to simplify the process in order to reduce the capital cost of a lignocellulose-to-ethanol facility. With this in mind, five research advances were identified that are required for process simplification:

1. development of biocatalysts with improved resistance to hemicellulose toxins (eliminates the need for separate detoxification steps);
2. replacement of sulfuric acid with the less aggressive phosphoric acid (eliminates the need for expensive metals or alloys);
3. solving the mixing and pumping issues related to high fiber solids loading (simplifies material handling, reduces opportunities for contamination, and improves product yields);
4. limiting the use of chemicals to those that are nutrients for the biocatalyst and for ultimate use as a high nitrogen fertilizer (partial recovery of chemical cost through multiple usage);
5. co-fermentation of hexose and pentose sugars in the same vessel (eliminates early liquid solid separation, fiber washing and detoxification of hemicellulose hydrolysate).

In recent years, significant progress has been made in this area with the development of a process termed Liquefaction plus Simultaneous Saccharification and co-Fermentation (L + SScF) that is analogous to corn ethanol (Geddes et al., 2011). This process uses a dilute-phosphoric-acid-steam-explosion pretreatment and meets target criteria for process simplification. In addition, the engineered *Escherichia coli* as the microbial biocatalyst is able to co-ferment all sugars derived from the lignocellulosic biomass while maintaining high conversion yields (Geddes et al., 2013). The L + SScF process was successfully scaled up to 80-L using a unit operation pilot plant (Nieves et al., 2011). Information gained from these studies was used to design the Stan Mayfield Biorefinery, a state-funded facility with

the purpose of proving the feasibility of this lignocellulosic ethanol technology in a larger, fully integrated, continuous process that resembles a commercial-scale plant.

In this present study, data from the biorefinery pilot plant, and laboratories were used to develop a techno-economic model for the construction of a 83 million liters per year (22 million gallons of ethanol per year) commercial plant in order to determine the economic feasibility of the process and to identify areas for further improvement. An experimentally proven case and 5 hypothetical scenarios were evaluated in which enzyme loading, enzymatic glucan hydrolysis, overall biomass-to-ethanol conversion, solids loading, and incubation time are varied. Scenarios were also compared in terms of heat demand, electricity, fertilizer production, and cost of ethanol production.

2. Methods

2.1. Bagasse to ethanol plant

The proposed ethanol plant is assumed to be located in the United States and to convert 300,000 dry US tons of sugarcane bagasse into ethanol annually. It is assumed to be in operation for 8000 h/year. Live steam is assumed to be available at 20 and 4 bar. Whenever possible, secondary steam is used to replace live steam. The process model was obtained by modifying previous models reported by Barta et al. (2010) using a process design based on Nieves et al. (2011). Description of the process steps will focus primarily on modifications made to the model.

2.1.1. Feedstock

The dry matter (DM) of sugarcane bagasse contains approximately 43% glucan, 23% xylan, 2% arabinan, 2% galactan and 27% lignin (Nieves et al., 2011). The remaining portion is acetyl groups, ash and other compounds. The moisture content of bagasse as received is typically 50%.

2.1.2. Pretreatment

In the modelled process (Fig. 1) the conversion of carbohydrates is carried out after dilute-acid steam explosion pretreatment and in L + SScF. Additional details of compositions, energy, flows, etc. are provided as Supplemental tables and figures. The sugarcane bagasse is pretreated using dilute phosphoric acid (the conversion factors for some reactions are the following: glucan to glucose 0.021, xylan to xylose 0.727, xylan to furfural 0.099, arabinan to furfural 0.500, water-insoluble lignin to water-soluble lignin 0.148), after which a small part of the liquid fraction is separated from the pretreated biomass slurry (30% dry weight; DW) and used for seed propagation (ethanologenic *E. coli* SL100).

2.1.3. Enzymatic liquefaction and ethanol fermentation

The liquid fraction (6.2% of the DW flow from the pretreatment reactor, 10% DW content) is used for propagating the fermenting organism *E. coli* SL100. This genetically modified strain ferments both hexoses and pentoses into ethanol. The solid fraction (34% DW) is liquefied by Novozymes (Franklinton, NC, USA) CTec3® cellulase enzyme-preparation (50 °C, 6 h, enzyme loading: 2.5% of the DW in the Base Case) and then Simultaneous Saccharification and co-Fermentation is carried out (pH 6.3, 37 °C, 48 h in the Base Case), in which both the hexoses and pentoses are fermented into ethanol (conversion factors for some reactions in the Base Case: glucan to glucose 0.68, galactan to galactose 1.000, xylan to xylose 0.500, glucose to ethanol 0.950, galactose to ethanol 0.950, xylose to ethanol 0.900). For pH control during liquefaction and fermentation, ammonia is used. The *E. coli*-concentration in the fermentation broth is 2.6% (w/w), and the solids loading for the Base Case was 15% (w/w).
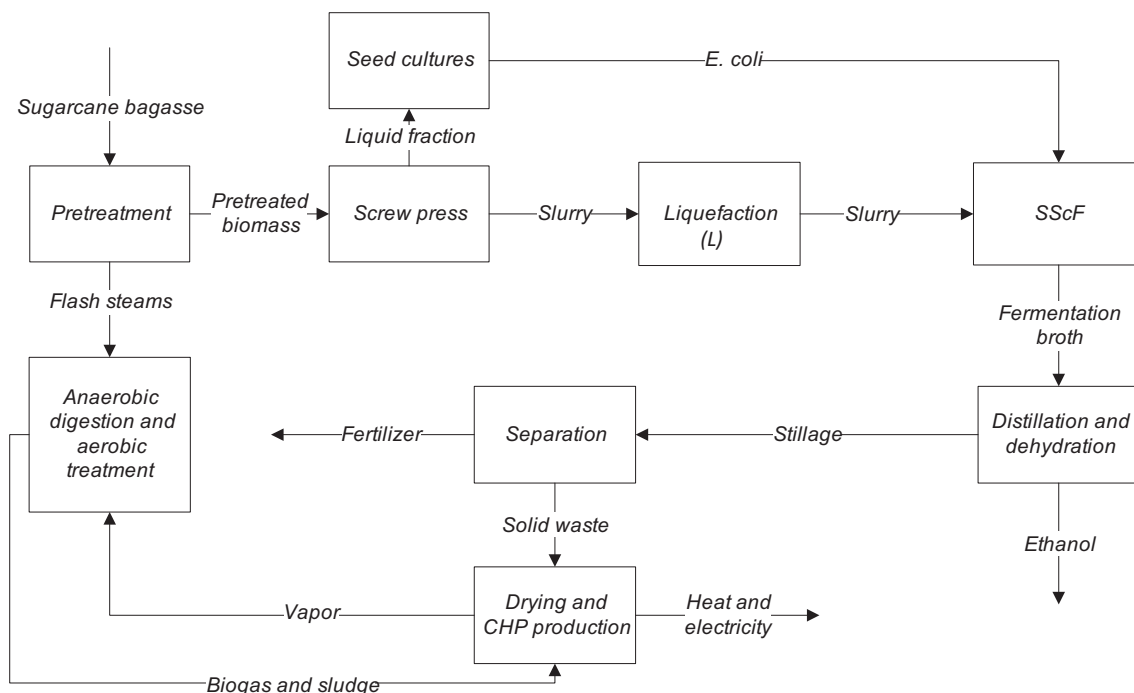
Fig. 1. Process scheme. After dilute acid pretreatment, approximately 6% of the solids were separated using a screw press. Hydrolysate liquid (10% dry weight) was used as a source of sugars for seed production. The remaining 94% is not separated and flows directly into the liquefaction tank. All materials (seed cultures and hydrolysate slurry) were combined in the main fermentation tanks. *Abbreviations:* CHP, combined heat and power; L + SScF, Liquefaction plus Simultaneous Saccharification and co-Fermentation.

2.1.4. Distillation and dehydration

The ethanol is purified in distillation and dehydration (molecular sieve adsorption) steps in order to produce pure (99.8% w/w) ethanol. The distillation step consists of two parallel operating stripper columns (25 trays) to separate the ethanol from the fermentation broth and a rectifier (45 trays) to concentrate the ethanol to 94%, which are heat integrated by operating at different pressures in order to reduce the energy demand. Ethanol recovery is assumed to be 99.5% in each column.

2.1.5. Separation and drying

The stillage is separated into liquid and solid fractions in a filter press. The liquid fraction is sold as fertilizer. The solid fraction is dried in a rotary drum drier operating at atmospheric pressure. The flue gases from the combined heat and power (CHP) unit, and live steam are used in the drying process as heating media. The flue gases are mixed with the solid fraction (direct heating), and then cool down from 150 °C to 105 °C. Flash steam from pretreatment is used to heat the surface of the dryer (indirect heating).

2.1.6. Anaerobic digestion and aerobic wastewater treatment

The condensed flash streams originating from pretreatment and drying are treated by AD (anaerobic digestion) followed by an aerobic treatment step. Anaerobic digestion is performed under mesophilic conditions and, hence, the inlet flow is cooled down to 37 °C before being fed to the first digester. The methane and AD sludge yields are assumed to be 0.35 Nm³/kg COD (chemical oxygen demand) removed and 0.03 kg sludge DM/kg COD fed, respectively. The effluent from AD is treated aerobically. The organic matter is removed almost entirely and sludge is produced at a yield of 0.3 kg sludge DM/kg organic matter.

2.1.7. Power production

Steam and electricity are generated by burning the solid fraction of the stillage, the biogas, and the sludge. The generated superheated steam is allowed to expand to 4 bar through a high-

pressure turbine system. Part of the steam is withdrawn at 20 bar for pretreatment. The produced excess heat is used to generate electricity through an additional low-pressure turbine with a discharge pressure of 0.75 bar.

2.2. Scenarios investigated

The important input details of each scenario are given in Table 1. One scenario is based on the experimental data obtained in a pilot scale run in the Stan Mayfield Biorefinery. The time of fermentation, the enzyme loading, the enzymatic glucan conversion, the overall biomass-to-ethanol conversion, and the solids loading are varied arbitrarily to investigate different enzyme loading strategies. The Base Case and Scenario 2 are only different in the time of SScF. For Scenarios 3 and 4, the parameters used represent projected values that were considered to be attainable as the technology matures, while the rest of the scenarios use values that can be obtained with current technology.

2.3. Process design and economics

Mass and energy balances were solved using the commercial flow sheeting software Aspen Plus V8.0 (Aspen Technology, Inc.,

Table 1
Input parameters of the various scenarios using L + SScF process. (Liquefaction plus Simultaneous Saccharification and co-Fermentation).

Scenario	Exp.	BC	1	2	3	4
Time of SScF (after liquefaction), h	72	48	48	36	36	36
Enzyme loading, % of dry weight	5	2.5	5	2.5	1.25	1.25
Enzymatic glucan conversion, % of theoretical	0.68	0.68	0.83	0.68	0.68	0.74
Solids loading, %	12	16	16	16	19	16
Overall process yield, g EtOH/g dry biomass	0.241	0.241	0.276	0.241	0.241	0.255

L + SScF, Liquefaction plus Simultaneous Saccharification and co-Fermentation; experimental parameters, Exp.; Base Case, BC.

Cambridge, MS, USA). This software is able to solve mass and energy balances and calculate the thermodynamic properties of all streams in the process. Data for the physical properties were obtained from the built-in database of Aspen Plus, or from the NREL database for biomass components such as polysaccharides and lignin. Aspen Energy Analyzer V8.0 (Aspen Technology, Inc.) was used to design a near-optimal heat exchanger network by means of Pinch technology and to estimate its capital cost. The requirements for heating and cooling capacity were fed back to the process model in Aspen Plus.

The fixed capital investment cost (excluding the heat exchanger network) was estimated either with the Aspen Process Economic Analyzer V8.0 (Aspen Technology, Inc.) setting 2012 as costing year, or from vendor quotations (in the cases of the pretreatment unit, filter presses, dehydration system, CSTR anaerobic digesters with their feed systems, steam boiler, and the flue gas condenser). The annualized fixed capital cost was determined by multiplying the fixed capital investment by an annuity factor of 0.11, corresponding to a 15-year depreciation period and an interest rate of 7%. The annual working capital is the product of the working capital investment and the interest rate. Table 2 summarizes the cost of operation and purchase prices for materials. The ethanol production cost was calculated as follows:

in the glucan conversion. The on-site *E. coli*-propagation has the advantage that pentoses, not fermentable by ordinary baker's yeast, can be utilized during this step. The amount of *E. coli* needed for the fermentation is the same in all the scenarios, therefore the amount of sugars consumed during the propagation step was also the same in each case. As the overall enzymatic glucan conversion is the highest in Scenario 1 and the mass flow of pentoses fed into L + SScF is the same in all cases, this scenario has the highest ethanol concentration after L + SScF. The ethanol concentrations in the L + SScF broth varied between 3.4% and 5.15% (w/w).

After heat integration the overall heat duty of the process varies in a narrow range, hence the significant differences in the specific heat demand are due to the produced amount of ethanol. The inlet enthalpy flow of CHP is higher when the overall ethanol yield is lower. Since the heat duties after heat integration are similar in the scenarios, more electricity is generated at higher inlet enthalpy flow of CHP, i.e. at lower overall ethanol yield. The power required by the process does not change significantly in the scenarios. While the power exported is 2.3–3.4-times of that required by the process, the excess heat duty used for power generation is 0.5–0.9 times of that required by the process after heat integration.

$$\text{Ethanol production cost} = \frac{\text{Annualized capital cost} + \text{Annual operating cost} - \text{Annual coproduct income}}{\text{Annual ethanol production}}$$

The plant is assumed to be located next to a sugar mill, with the feedstock already collected and readily available. Assuming a tonne of sugarcane generates 125 kg of dry bagasse and 85 kWh of excess electrical power (Pippo and Luengo, 2013) sold to the grid at 0.065 US$/kWh (Kim et al., 2013), the cost of the biomass would be 44 US$ per dry tonne bagasse. For the fertilizer, a 50% recovery of the cost of chemicals is assumed for the Base Case. The enzyme cost was back-calculated to obtain a value similar to those reported in the literature in terms of cost of enzyme per liter of ethanol produced (Chovau et al., 2013). The electricity has a conservative price, considering that the average US residential retail price in March 2012 was 11.76 cents per kWh (http://www.eia.gov/cneaf/electricity/epm/epm_sum.html). The rest of the prices used were from direct quotes from vendors.

3. Results and discussion

3.1. Mass and energy balance

Table 3 lists the components in the liquefaction step and in the SScF in each scenario. After the steam explosion pretreatment the cases are different in the amount of enzyme added, and therefore

Table 2
Prices associated with operational costs.

Feedstock	Sugarcane bagasse (tonne; bone dry)	44.00
Product income	Electricity (kWh)	0.04
	Fertilizer (L)	0.0045
Chemicals	H_3PO_4 (85%) (kg)	0.80
	NH_4OH (kg)	0.68
	$MgSO_4$ (kg)	0.30
	Sodium metabisulfite (kg)	0.40
	Cellulase (kg)	1.00
Utilities	Cooling water (m³)	0.02
	Process water (m³)	0.20
Other costs	Labor (per employee/year)	60,000.00

All prices are in US dollars.

3.2. Capital investment

The direct costs of pretreatment, L + SScF, and CHP are the largest contributors to the capital investment (Table 4). The L + SScF has a significantly higher cost when the fermentation time is 72 h, instead of 48 h or 36 h. Therefore, the highest cost for L + SScF was obtained in the experimental scenario, where it constituted

Table 3
Process details of the various scenarios based on the solved mass and energy balances.

Scenario	Exp.	BC and 2	1	3	4
Components into liquefaction, kg/h					
Hexosans	15,091	15,091	15,091	15,091	15,091
Pentosans	577	577	577	577	577
Hexoses	281	281	281	281	281
Pentoses	5702	5702	5702	5702	5702
CTec3 ® enzyme[1]	308	154	308	77	77
Components into SScF, kg/h					
Hexosans	6107	6107	4206	6107	5367
Pentosans	412	412	412	412	412
Hexoses	10,262	10,262	12,375	10,262	11,086
Pentoses	5889	5889	5889	5889	5889
E. coli	626	626	626	626	626
Ethanol at the end of SScF,% (w/w)	3.4	4.24	4.84	5.15	4.50
Overall heat duty after heat integration, kW	50,417	48,759	49,991	48,869	48,169
Inlet enthalpy flow of fuel to CHP[2], kW	87,952	74,486	77,062	73,457	73,336
Specific heat demand, MJ/L EtOH	17	17	15	19	16
Power produced, kW	11,750	11,105	11,585	11,148	10,876
Power required by the process, kW	3522	5026	4962	4975	4898
Power exported, kW	8228	6079	6623	6173	5978

Abbreviations: Exp., experimental parameters; BC, Base Case; L + SScF, Liquefaction plus Simultaneous Saccharification and co-Fermentation; CHP, combined heat and power.
[1] In the enzyme preparation 20% protein content was assumed.
[2] Based on the lower heating value of the fuel mixture.

Table 4
Breakdown of the total capital investment cost in million dollars (US).

Scenario	Exp.	BC	1	2	3	4
Raw material handling	2.64	2.64	2.64	2.64	2.64	2.64
Pretreatment	21.90	21.90	21.90	21.90	21.90	21.90
L + SScF	36.36	28.81	28.81	25.08	25.08	25.01
Distillation	4.80	4.62	4.62	4.62	4.48	4.57
PneumaPress	9.22	13.13	13.13	13.13	13.13	13.13
Drying	4.31	4.31	4.31	4.31	4.31	4.31
CHP	31.30	31.28	31.28	31.19	31.19	31.19
Storage	2.99	2.99	2.99	2.99	2.99	2.99
Heat exchanger network	2.48	2.32	1.97	2.32	2.25	2.20
WWT	7.97	8.09	8.06	8.09	8.22	8.08
Total direct cost	123.98	120.09	119.71	116.27	116.20	116.02
Total indirect cost	66.55	54.47	54.07	47.16	46.95	46.89
Fixed capital investment	190.53	174.56	173.78	163.43	163.15	162.91
Working capital	6.29	6.12	6.87	6.12	6.12	6.42
Total capital investment	196.82	180.68	180.65	169.55	169.28	169.33

Abbreviations: Exp., experimental parameters; BC, Base Case; L + SScF, Liquefaction plus Simultaneous Saccharification and co-Fermentation; CHP, combined heat and power; WWT, wastewater treatment.

29% of the total direct cost (versus 21–24% for the other scenarios). This scenario requires the highest fixed capital investment, while those of the Base Case and Scenarios 1–4 are significantly lower, and vary in a narrow range.

In the Base Case, savings of ~10 million US$ have been estimated for the phosphoric acid pretreatment device which does not require construction from exotic alloys or zirconium. This translates to ~6% savings in the fixed capital investment without taking into consideration other savings such as the elimination of liquid solid separation steps and gypsum handling. Additional savings may be obtained by using a continuous fermentation configuration instead of a batch system. This will reduce the fermentation down-time considerably, thereby reducing the amount of fermentation vessels necessary and the capital investment cost of the L + SScF process. In addition, the amount of vessels required for the seed train would be significantly less, reducing further the capital cost and introducing additional savings in the operational costs.

3.3. Ethanol production costs

The cost elements of ethanol production expressed in US cents/L ethanol are listed in Table 5. In all the scenarios the feedstock and the capital costs were found to be the main cost contributors. The feedstock and the capital cost constitute between 23–28% and 40–49% of the total cost, respectively. The cost of utilities (process and

Table 5
Breakdown of the ethanol production cost (US cent/L ethanol).

Scenario	Exp.	BC	1	2	3	4
Cost, US cent/L EtOH						
Feedstock	14.47	14.47	12.69	14.47	14.47	13.70
Capital	27.66	26.24	23.00	25.46	25.45	24.09
Chemicals	7.91	7.91	6.94	7.91	7.91	7.49
Enzymes	14.77	7.38	12.95	7.38	3.69	3.50
Utilities	0.88	0.75	0.68	0.75	0.65	0.70
Other	7.84	7.65	6.70	7.55	7.55	7.15
Co-product income, US cent/L EtOH						
Fertilizer	7.65	3.95	3.47	3.95	3.95	3.74
Electricity	3.16	2.58	2.28	2.62	3.95	2.49
Ethanol production cost, US cent/L EtOH	62.72	57.87	57.19	56.93	51.80	50.38

Other refers to maintenance, insurance and labor.
Abbreviations: EtOH, ethanol; Exp, experimental values; BC, Base Case.

cooling water) is negligible, since steam and electricity demands of the process are covered by on-site steam and power generation.

The experimental case yielded the highest ethanol production cost. This can be explained by the longer fermentation time (resulting in increased capital investment) and the lower solids loading with high enzyme dosage. On the other hand, increasing the ethanol yield by 14.5% (from 0.241 in the Base Case to 0.276 in Scenario 1) results in a lower ethanol production cost, even with a higher enzyme loading. Hence, it is a good strategy to focus on obtaining the highest overall ethanol yield possible, and then optimize the unit operations to reduce costs without lowering the overall yield. This is based on assumptions made regarding the cost of commercial enzymes. However, the literature values for enzyme cost vary considerably, with some reporting minimal contribution to the production cost, while others identify the enzyme cost as a major hurdle to overcome (Klein-Marcuschamer et al., 2012). Based on some of the higher enzyme prices reported in the literature (Klein-Marcuschamer et al., 2012), the enzyme cost would be close to 30 US cents/L, making the enzyme cost the main contributor to the ethanol production cost. However, Novozymes reported in 2010 an enzyme cost of 13.2 US cents/L ethanol produced (Chovau et al., 2013; Kumar and Murthy, 2011), a target that may be achieved by improvements of commercial enzymes.

The biocatalyst used for this study is one capable of metabolizing hexose and pentose sugars with high metabolic yields. In addition, the biocatalyst also holds an increased tolerance to the metabolic inhibitors that are present as a result of sugar degradation during pretreatment (Geddes et al., 2013). These characteristics make possible the elimination of liquid/solid separation steps that are part of the SHF (Separate Hydrolysis and Fermentation), SSF (Simultaneous Saccharification and Fermentation), and SHcF (Separate Hydrolysis and co-Fermentation) processes that are usually followed by the detoxification of the solubilized hemicellulose fraction (Foust et al., 2009; Kim et al., 2013; Kumar and Murthy, 2011; Sarkar et al., 2012).

To assess the impact of these process simplifications in the MESP, two additional variations of the model were run. In the first one, a liquid/solid separation step was added after pretreatment, followed by a detoxification step for the hemicellulose sugars and the subsequent fermentation of cellulose and hemicellulose sugars together in a single fermentation step (Separate Hydrolysis and co-Fermentation; SHcF). In this process, there is an increase in the capital cost due to the additional equipment necessary for the added separation step, as well as for the detoxification (Fig. 2). In addition, the cost of chemicals is increased due to the chemicals
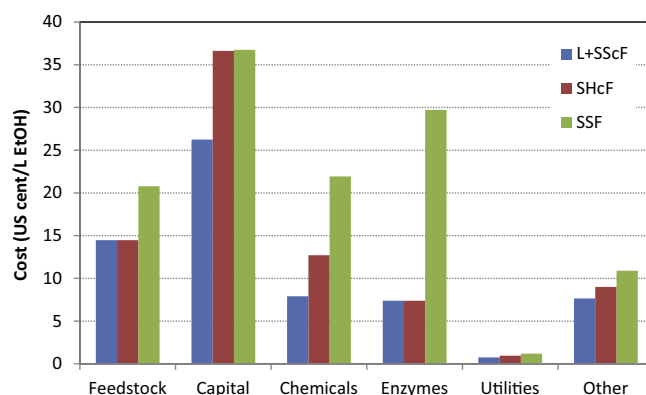

Fig. 2. Summary of the ethanol production cost for three different process configurations. Liquefaction plus Simultaneous Saccharification and co-Fermentation (L + SScF), Separate Hydrolysis and co-Fermentation (SHcF), and Simultaneous Saccharification and Fermentation (SSF). Other refers to maintenance, insurance, and labor.
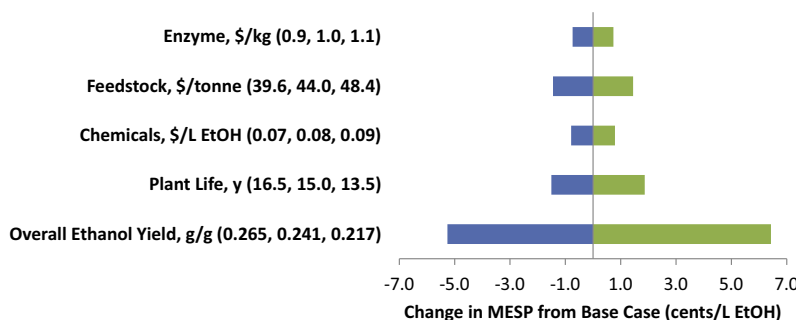
Fig. 3. Sensitivity of the ethanol production cost to a 10% change in several parameters from the values used in the Base Case. Listed is the name of the parameter, the units of the parameter, and in parenthesis the three values used for the sensitivity analysis. The overall ethanol yield is expressed as the grams of ethanol produced per gram of initial biomass processed on a dry weight basis.

needed for the detoxification of the hemicellulose sugars. The other process configuration modelled was the Simultaneous Saccharification and Fermentation (SSF) process. This process consists of a liquid/solid separation after pretreatment, followed by enzyme hydrolysis and fermentation of the cellulose fraction. The SSF configuration had the highest production cost of all three process configurations studied due to the lower overall ethanol yields that are obtained when the hemicellulose sugars are not used (Fig. 2).

The co-products generated in the process are electricity and fertilizer, and both of them are assumed to be marketable. Scenario 1 has the lowest electricity excess due to the high ethanol yield, and Scenario 3 has the highest amount of sold electricity due to the lower ethanol yield and higher solids loading. A higher solids loading seems to be beneficial, but not as much as a higher ethanol yield. The ethanol production cost varies between 50.38 and 62.72 US cents/L, with the experimental case increasing to more than 70 cents/L without fertilizer income. The use of the remaining liquid at the end of the process as fertilizer is made possible by limiting the use of chemicals to those that are nutrients for the biocatalyst and can be used as a high nitrogen fertilizer (ammonia for pH adjustment and phosphoric acid during pretreatment). The ethanol production cost would increase 6.8% (from 57.87 to 61.82 US cents/ L ethanol) in the base case if the sale of the fertilizer was not possible.

Fig. 3 shows how the ethanol production cost changes, in terms of US cents/L ethanol produced, when there is a ±10% variation in the different parameters. The largest effect by far is the overall ethanol yield. A 10% increase in the ethanol yield results in savings of more than 5 US cents/L (MESP of 52.61 US cents/L), while a decrease of 10% increased the production cost more than 6 US cents/L to 64.3. This is in agreement with the need to focus on obtaining the highest overall ethanol yield, even if a higher enzyme dosage is used. The feedstock price and the plant life have a similar impact in the production cost, while the enzyme cost and the price of chemicals have the lowest effect.

3.4. Comparison to NREL model

As stated earlier in the introduction, due to the variability of assumptions and process configurations used between different models, it is difficult to make a direct comparison between models that have been presented by different groups. In addition, the information needed to understand the underlying assumptions of the models is often not included, complicating matters further. However, the NREL reports have served as a benchmark for other technoeconomic models published (Humbird et al., 2011) and as a means to show the potential of the UF phosphoric process.

Table 6 lists several of the main parameters that were used to build the 2011 NREL model and this study's model. The largest dis-

Table 6
Parameters used to generate the technoeconomic model and the values assumed by the NREL 2011 study (Humbird et al., 2011) and the current analysis. Dollar values are expressed in 2007 dollars. TCI, total capital investment; IRR, Internal Rate of Return.

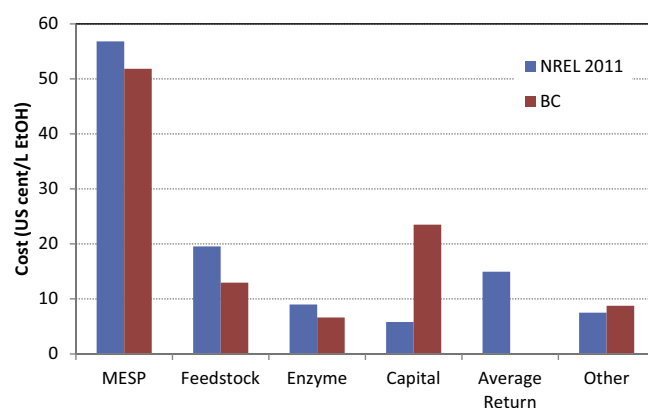| Parameter	NREL report	BC
Tonnes/day	2000	820
L/tonne	329	305
US$/tonne	64.90	39.41
Overall yield	0.260	0.241
ML/year	231	83
TCI (mil US dollars)	422.5	161.81
TCI/annual liter	1.83	1.95
Percentage equity financed	40%	100%
Plant life (years)	30	15
After-tax IRR	10%	0%
Operating hours/year	8410	8000

Fig. 4. Comparison of the production costs between the 2011 NREL report (Humbird et al., 2011) and the BC of the current study. The values are expressed in 2007 dollars. Average Return refers to the average return on investment for the NREL model assuming 10% after-tax Internal Rate of Return (IRR) and 40% equity financing with a 10 year loan at 8% interest. Other includes chemicals, co-products sold (those would be a negative cost), maintenance, insurance, and labor. MESP, minimum ethanol selling price.

crepancies in the assumptions are based on the scale that was used for the model. A larger plant capacity would take advantage of the scale to reduce the production cost. The current model uses a size that is closer to the current state of technology. In addition, the plant life is assigned 15 years, while the NREL model used 30 years. This difference in the capital cost and plant life, as is reflected in the MESP, is displayed in Fig. 4. Fig. 4 displays a side-by-side comparison of the production cost for the 2011 NREL and the current study BC models with the values expressed in 2007 dollars.

Another difference between the models is on the equity financing and the Internal Rate of Return (IRR). The UF phosphoric acid model assumes an IRR equal to zero with 100% equity financing, while the NREL model assumed an after-tax IRR of 10% and 40% equity financing. These differences adding an additional 15 cents per liter to the production cost of the NREL model (Fig. 4). On the other hand, the cost of the feedstock is 50.8% higher for the NREL model. As explained in the methods section, feedstock were assumed to be collected and ready to use. Price was calculated by considering money that the sugar mill would lose by not burning the biomass for power and instead allowing the use in the biorefinery.

It is interesting to note that the enzyme price is 36% higher in the NREL model. It seems that the continued improvement of commercial lignocellulosic enzymes will continue to drive the price down, in which case it may make more financial sense to purchase the enzymes than to produce on site. Overall, the MESP is 9.6% higher for the NREL model, indicating the financial viability of the UF phosphoric acid based process.

4. Conclusions

The technology developed at the Stan Mayfield Biorefinery Pilot Plant in which sugarcane bagasse is converted into ethanol, fertilizer and electricity using a genetically modified *E. coli* strain that ferments both hexoses and pentoses, has been evaluated in a techno-economic analysis. The identified main cost contributors are the feedstock price and the annualized capital cost, which contribute 25% and 45% of the total ethanol production cost, respectively. In addition, the ethanol yield has an important effect on the production cost, and as per the model it is financially acceptable to use higher enzyme concentrations to increase the ethanol yield.

Acknowledgements

Appendix A. Supplementary data

Supplementary data associated with this article can be found, in the online version, at http://dx.doi.org/10.1016/j.biortech.2016.01.093.

References

Aden, A., Foust, T., 2009. Technoeconomic analysis of the dilute sulfuric acid and enzymatic hydrolysis process for the conversion of corn stover to ethanol. Cellulose 16, 535–545.

Aden, A., Ruth, M., Ibsen, K., Jechura, J., Neeves, K., Sheehan, J., Wallace, B., Montague, L., Slayton, A., and Lukas, J. 2002. Lignocellulosic biomass to ethanol process design and economics utilizing co-current dilute acid prehydrolysis and enzymatic hydrolysis for corn stover. Golden (CO): National Renewable Energy Laboratory. Report no.: NREL/TP-510-32438. Contract no.: DE-AC36-99-GO10337.

Barta, Z., Reczey, K., Zacchi, G., 2010. Techno-economic evaluation of stillage treatment with anaerobic digestion in a softwood-to-ethanol process. Biotechnol. Biofuels 3, 21.

Chovau, S., Degrauwe, D., van der Bruggen, B., 2013. Critical analysis of techno-economic estimates for the production cost of lignocellulosic bio-ethanol. Renew. Sustain. Energy Rev. 26, 307–321.

Eggeman, T., Elander, R.T., 2005. Process and economic analysis of pretreatment technologies. Bioresour. Technol. 96, 2019–2025.

Foust, T.D., Aden, A., Dutta, A., Phillips, S., 2009. An economic and environmental comparison of a biochemical and a thermochemical lignocellulosic ethanol conversion process. Cellulose 16, 547–565.

Galbe, M., Sassner, P., Wingren, A., Zacchi, G., 2007. Process engineering economics of bioethanol production. Adv. Biochem. Eng. Biotechnol. 108, 303–327.

Geddes, C.C., Mullinnix, M.T., Nieves, I.U., Peterson, J.J., Hoffman, R.W., York, S.W., Yomano, L.P., Miller, E.N., Shanmugam, K.T., Ingram, L.O., 2011. Simplified process for ethanol production from sugarcane bagasse. Bioresour. Technol. 102, 2702–2711.

Geddes, C.C., Mullinnix, M.T., Nieves, I.U., Hoffman, R.W., Sagues, W.J., York, S.W., Shanmugam, K.T., Erickson, J.E., Vermerris, W.E., Ingram, L.O., 2013. Seed train development for the fermentation of bagasse from sweet sorghum and sugarcane using a simplified fermentation process. Bioresour. Technol. 128, 716–724.

Hamelinck, C.N., van Hooijdonk, G., Faaij, A.P.C., 2005. Ethanol from lignocellulosic biomass: techno-economic performance in short-, middle- and long-term. Biomass Bioenergy 28, 384–410.

Han, J., Luterbacher, J.S., Alonso, D.M., Dumesic, J.A., Maravelias, C.T., 2015. A lignocellulosic ethanol strategy via nonenzymatic sugar production: process synthesis and analysis. Bioresour. Technol. 182, 258–266.

Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., Aden, A., Schoen, P., Lukas, J., Olthof, B., Worley, M., Sexton, D., and Dudgeon, D. 2011. Process design and economics for biochemical conversion of lignocellulosic biomass to ethanol – dilute-acid pretreatment and enzymatic hydrolysis of corn stover. Golden (CO): National Renewable Energy Laboratory. Report no.: NREL/TP-5100-47764. Contract no.: DE-AC36-08GO28308.

Kim, J., Sen, S.M., Maravelias, C.T., 2013. An optimization-based assessment framework for biomass-to-fuel conversion strategies. Energy Environ. Sci. 6, 1093–1104.

Klein-Marcuschamer, D., Oleskowics-Popiel, P., Simmons, B.A., Blanch, A.W., 2012. The challenge of enzyme cost in the production of lignocellulosic biofuels. Biotechnol. Bioeng. 109, 1083–1087.

Kumar, D., Murthy, G.S., 2011. Impact of pretreatment and downstream processing technologies on economics and energy in cellulosic ethanol production. Biotechnol. Biofuels 4, 27.

Lynd, L.R., Cushman, J.H., Nichols, R.J., Wyman, C.E., 1991. Fuel ethanol from cellulosic biomass. Science 251, 1318–1323.

Macrelli, S., Mogensen, J., and Zacchi, G. 2012. Techno-economic evaluation of 2nd generation bioethanol production from sugar cane bagasse and leaves integrated with the sugar-based ethanol process.

Nguyen, Q.A., Saddler, J.N., 1991. An integrated model for the technical and economic evaluation of an enzymatic biomass conversion process. Bioresour. Technol. 35, 275–282.

Nieves, I.U., Geddes, C.C., Mullinix, M.T., Hoffman, R.W., Tong, Z., Castro, E., Shanmugam, K.T., Ingram, L.O., 2011. Injection of air into the headspace improves fermentation of phosphoric acid pretreated sugarcane bagasse by Escherichia coli MM170. Biores. Technol. 102, 6959–6965.

Pippo, W.A., Luengo, C.A., 2013. Sugarcane energy use: accounting of feedstock energy considering current agro-industrial trends and their feasibility. Int. J. Energy Environ. Eng. 4, 10.

Sassner, P., Galbe, M., Zacchi, G., 2008. Techno-economic evaluation of bioethanol production from three different lignocellulosic materials. Biomass Bioenergy 3, 422–430.

Sarkar, N., Ghosh, S.K., Bannerjee, S., Aikat, K., 2012. Bioethanol production from agricultural wastes: an overview. Renew. Energy 37, 19–27.

von Sivers, M., Zacchi, G., 1995. A techno-economical comparison of three processes for the production of ethanol from pine. Biores. Technol. 51, 43–52.

Wooley, R., Ruth, M., Sheehan, J., Ibsen, K., Majdeski, H., and Galvez, A. 1999. Lignocellulosic biomass to ethanol process design and economics utilizing co-current dilute acid prehydrolysis and enzymatic hydrolysis current and futuristic scenarios. Golden (CO): National Renewable Energy Laboratory. Report no.: NREL/TP-580-26157. Contract no.: DE-AC36-98-GO10337.

Wright, M.M., Satrio, J.A., Brown, R.C., Daugaard, D.E., Hsu D.D., Techno-economic analysis of biomass fast pyrolysis to transportation fuels, NREL Technical Report, 2010.

Wyman, C.E., 1994. Ethanol from lignocellulosic biomass: technology, economics, and opportunities. Biores. Technol. 50, 3–16.