

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

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12 **Running Head:** Techno-economic analysis of L+SScF process for ethanol

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14 **For Consideration by:** Bioresource Technology Journal

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22 **Key Words:** lignocellulose, xylose, ethanol, hemicellulose, techno-economic analysis,  
23 L+SScF

## **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 time of L+SScF, 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 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 Sievers 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; Kumar and Murthy, 2011; Macrelli et al, 2012). These differences have made it difficult to compare these studies (Chovau et al, 2013; Galbe et al, 2007). The NREL report 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).

1           Some of the significant contributors to the MESP of lignocellulosic ethanol include the  
2 cost of the feedstock, the ethanol yield, and the cost of cellulase enzymes (Chovau et al.,  
3 2013). However, the main contributor of the MESP in almost all cases seems to be the capital  
4 cost (Galbe et al, 2007). From various studies, it is clear that one way to lower the MESP is  
5 to simplify the process in order to reduce the capital cost of a lignocellulose-to-ethanol  
6 facility. With this in mind, we identified five research advances that are required for process  
7 simplification:

- 8           1. Development of biocatalysts with improved resistance to hemicellulose toxins  
9           (eliminates the need for separate detoxification steps);
- 10          2. Replacement of sulfuric acid with the less aggressive phosphoric acid (eliminates  
11          the need for expensive metals or alloys);
- 12          3. Solving the mixing and pumping issues related to high fiber solids loading  
13          (simplifies material handling, reduces opportunities for contamination, and  
14          improves product yields);
- 15          4. Limiting the use of chemicals to those that are nutrients for the biocatalyst and for  
16          ultimate use as a high nitrogen fertilizer (partial recovery of chemical cost through  
17          multiple usage);
- 18          5. Co-fermentation of hexose and pentose sugars in the same vessel (eliminates early  
19          liquid solid separation, fiber washing and detoxification of hemicellulose  
20          hydrolysate).

21  
22           In recent years we have made significant progress in this area with the development of a  
23 process termed Liquefaction plus Simultaneous Saccharification and co-Fermentation (L+SScF) that is  
24 analogous to corn ethanol (Geddes et al, 2011). This process uses a dilute-phosphoric-acid-steam-  
25 explosion pretreatment and meets target criteria for process simplification. In addition, the engineered  
26 *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, the data from the biorefinery, pilot plant, and laboratory were used to develop a techno-economic model for production of 23 million gallons of ethanol per year (88 million liters 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. Materials and 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 8,000 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 (Figure 1) the conversion of carbohydrates is carried out after dilute-acid steam explosion pretreatment and in L+SScF. 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 *Escherichia 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 *Escherichia coli*. 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).

#### 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 Nm<sup>3</sup>/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

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

## 4 *2.2. Scenarios investigated*

5 The important input details of each scenario are given in Table 1. One scenario is  
6 based on the experimental data obtained in a pilot scale run in the Stan Mayfield Biorefinery.  
7 The time of fermentation, the enzyme loading, the enzymatic glucan conversion, the overall  
8 biomass-to-ethanol conversion, and the solids loading are varied arbitrarily to investigate  
9 different enzyme loading strategies. The Base Case and Scenario 2 are only different in the  
10 time of SS<sub>CF</sub>. For scenarios 3 and 4, the parameters used represent projected values that we  
11 consider to be attainable as the technology matures, while the rest of the scenarios use values  
12 that can be obtained with current technology.

## 13 *2.3. Process design and economics*

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

23 The fixed capital investment cost (excluding the heat exchanger network) was  
24 estimated either with the Aspen Process Economic Analyzer V8.0 (Aspen Technology, Inc.)  
25 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:

*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](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 SSsF in each scenario. After the steam explosion pretreatment the cases are different in the amount of enzyme added,

1 and therefore in the glucan conversion. The on-site *E. coli*-propagation has the advantage that  
2 pentoses, not fermentable by ordinary baker's yeast, can be utilized during this step. The  
3 amount of *E. coli* needed for the fermentation is the same in all the scenarios, therefore the  
4 amount of sugars consumed during the propagation step was also the same in each case. As  
5 the overall enzymatic glucan conversion is the highest in Scenario 1 and the mass flow of  
6 pentoses fed into L+SScF is the same in all cases, this scenario has the highest ethanol  
7 concentration after L+SScF. The ethanol concentrations in the L+SScF broth varied between  
8 3.4 and 5.15% (w/w).

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

### 17 3.2. Capital investment

18 The direct costs of pretreatment, L+SScF, and CHP are the largest contributors to the  
19 capital investment (Table 4). The L+SScF has a significantly higher cost when the  
20 fermentation time is 72 h, instead of 48 h or 36 h. Therefore, the highest cost for L+SScF was  
21 obtained in the experimental scenario, where it constituted 29% of the total direct cost (versus  
22 21% - 24% for the other scenarios). This scenario requires the highest fixed capital  
23 investment, while those of the Base Case and Scenarios 1-4 are significantly lower, and vary  
24 in a narrow range.

In the Base Case we have estimated savings of ~10 million US\$ for our pretreatment unit as a result of using phosphoric acid instead of sulfuric acid. 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 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

1 identify the enzyme cost as a major hurdle to overcome (Klein-Marcuschamer et al., 2012).  
2 Based on some of the higher enzyme prices reported in the literature (Klein-Marcuschamer et  
3 al., 2012), the enzyme cost would be close to 30 US cents/L, making the enzyme cost the  
4 main contributor to the ethanol production cost. However, Novozymes reported in 2010 an  
5 enzyme cost of 13.2 US cents/L ethanol produced (Chovau et al, 2013, Kumar and Murthy,  
6 2011), suggesting our reported values are realistic considering the significant improvements  
7 of commercial enzymes in the last few years.

8         The co-products generated in the process are electricity and fertilizer, and both of them  
9 are assumed to be marketable. Scenario 1 has the lowest electricity excess due to the high  
10 ethanol yield, and Scenario 3 has the highest amount of sold electricity due to the lower  
11 ethanol yield and higher solids loading. A higher solids loading seems to be beneficial, but  
12 not as much as a higher ethanol yield. The ethanol production cost varies between 50.38 and  
13 62.72 US cents/L, with the experimental case increasing to more than 70 cents/L without  
14 fertilizer income. The use of the remaining liquid at the end of the process as fertilizer is  
15 possible due to the use of phosphoric acid during pretreatment and the use of ammonia for pH  
16 adjustment. Further studies need to be completed in order to assess the feasibility of using  
17 this solution as fertilizer.

18         Another important thing to consider is the scale of the process that we have designed  
19 (Galbe et al, 2007). While other models used more than double the plant capacity of this  
20 study, thus taking advantage of the scale to reduce the production cost, we have developed our  
21 model using a size that is closer to the current state of technology. In addition, the plant life  
22 for our study is 15 years, which has an effect in the production cost, as demonstrated by the  
23 sensitivity analysis (Fig. 2).

24         Figure 2 shows how the ethanol production cost changes, in terms of US cents/L  
25 ethanol produced, when there is a  $\pm 10\%$  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.

#### **4. Conclusions**

The technology developed at the Stan Mayfield Biorefinery 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.

#### **5. Acknowledgments**

The authors gratefully acknowledge the Project OTKA PD-108389 of the Hungarian Research Fund and the New Hungary Development Plan (Project ID: TÁMOP-4.2.1/B-09/1/KMR-2010-0002) for their financial support. This research was also supported by the US Department of Agriculture (2011-10006-30358 and 2012-67009-19596), the US Department of Energy's Office of International Affairs (DE-PI0000031), BASF, and the Florida Department of Agriculture and Consumer Services.

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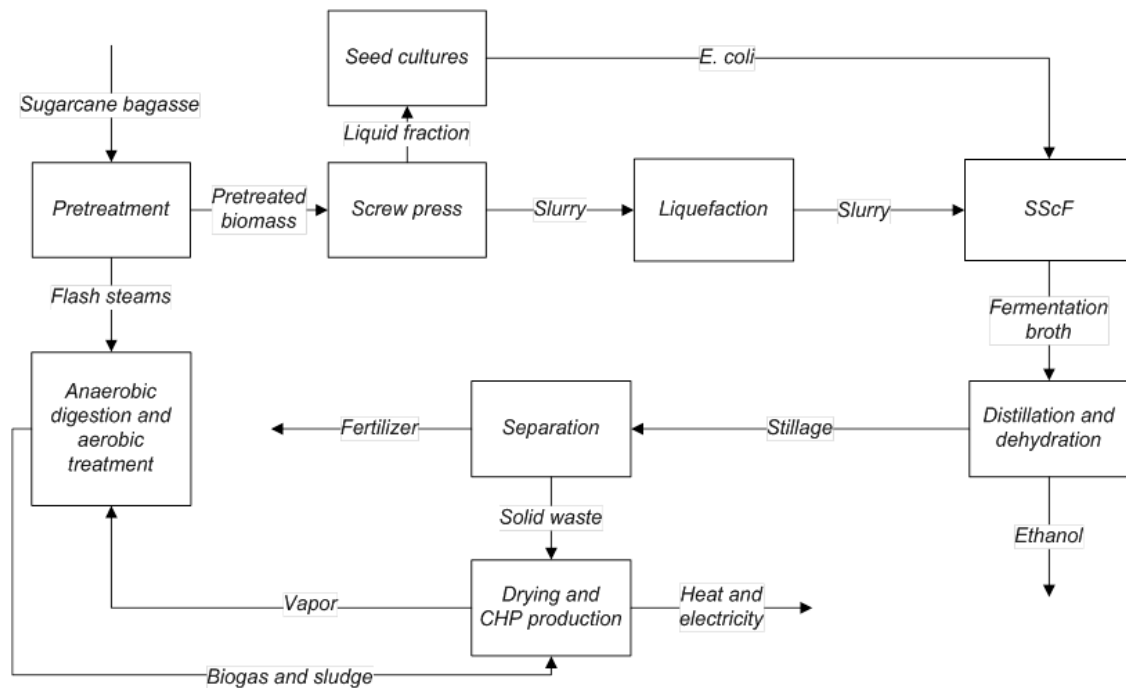
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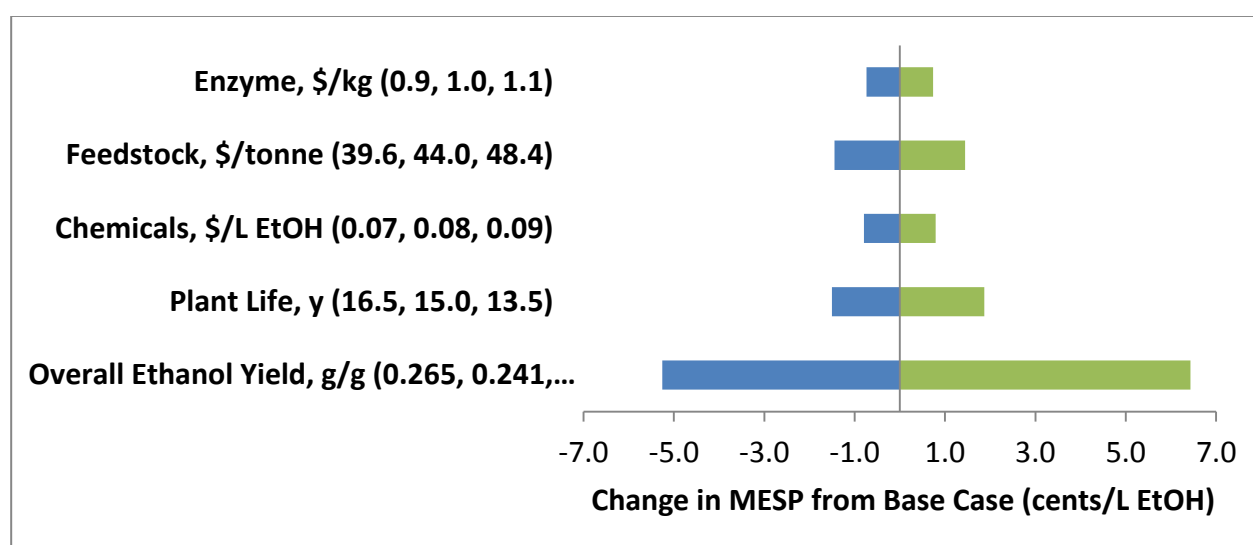


## Figure legends

**Figure 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.



**Figure 2.** 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.



1

2 **Table 1.** Input parameters of the various scenarios (BC, Base Case; SScF, simultaneous saccharification and co-fermentation).

Scenario	Exp.	BC	1	2	3	4
Time of SScF, 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

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1

2 **Table 2.** Prices associated with operational costs.

<b>Feedstock</b>	Sugarcane Bagasse (tonne; bone dry)	44.00
<b>Product income</b>	Electricity (kWh)	0.04
	Fertilizer (L)	0.0045
<b>Chemicals</b>	H <sub>3</sub> PO <sub>4</sub> (85%) (kg)	0.80
	NH <sub>4</sub> OH (kg)	0.68
	MgSO <sub>4</sub> (kg)	0.30
	Sodium Metabisulfite (kg)	0.40
	Cellulase (kg)	1.00
<b>Utilities</b>	Cooling water (m <sup>3</sup> )	0.02
	Process water (m <sup>3</sup> )	0.20
<b>Other costs</b>	Labor (per employee/year)	60,000.00

3 All prices are in US\$

4

1 **Table 3.** Process details of the various scenarios based on the solved mass and energy balances.  
2 Abbreviations: BC, Base Case; L+SScF, liquefaction plus simultaneous saccharification and co-  
3 fermentation; CHP, combined heat and power.

Scenario	Exp.	BC and 2	1	3	4
<i>Components into Liquefaction, kg/h</i>					
Hexosans	15091	15091	15091	15091	15091
Pentosans	577	577	577	577	577
Hexoses	281	281	281	281	281
Pentoses	5702	5702	5702	5702	5702
CTec3 ® enzyme <sup>1</sup>	308	154	308	77	77
<i>Components into SScF, kg/h</i>					
Hexosans	6107	6107	4206	6107	5367
Pentosans	412	412	412	412	412
Hexoses	10262	10262	12375	10262	11086
Pentoses	5889	5889	5889	5889	5889
<i>E.coli</i>	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	50417	48759	49991	48869	48169
Inlet enthalpy flow of fuel to CHP <sup>2</sup> , kW	87952	74486	77062	73457	73336
Specific heat demand, MJ/L EtOH	17	17	15	19	16
Power produced, kW	11750	11105	11585	11148	10876

Power required by the process, kW	3522	5026	4962	4975	4898
Power exported, kW	8228	6079	6623	6173	5978

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1 <sup>1</sup> In the enzyme preparation 20% protein content was assumed.

2 <sup>2</sup> Based on the lower heating value of the fuel mixture.

3

1 **Table 4.** Breakdown of the total capital investment cost in million USD. Abbreviations: BC,  
2 Base Case; L+SScF, liquefaction plus simultaneous saccharification and co-fermentation; CHP,  
3 combined heat and power; WWT, wastewater treatment.

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	<b>196.82</b>	<b>180.68</b>	<b>180.65</b>	<b>169.55</b>	<b>169.28</b>	<b>169.33</b>

**Table 5.** Breakdown of the ethanol production cost in US cent/L (Other refers to maintenance, insurance and labor). Abbreviations: Exp, experimental values; BC, Base Case.

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