

Techno-economic comparison of process technologies for biochemical ethanol production from corn stover[☆]

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ABSTRACT

This techno-economic study compares several process technologies for the production of ethanol from lignocellulosic material, based on a 5- to 8-year time frame for implementation. While several previous techno-economic studies have focused on future technology benchmarks, this study examines the short-term commercial viability of biochemical ethanol production. With that goal, yields (where possible) were based on publicly available experimental data rather than projected data. Four pretreatment technologies (dilute-acid, 2-stage dilute-acid, hot water, and ammonia fiber explosion or AFEX); and three downstream process variations (pervaporation, separate 5-carbon and 6-carbon sugars fermentation, and on-site enzyme production) were included in the analysis. Each of these scenarios was modeled and economic analysis was performed for an “*n*th plant” (a plant with the same technologies that have been employed in previous commercial plants) to estimate the total capital investment (TCI) and product value (PV). PV is the ethanol production cost, including a 10% return on investment. Sensitivity analysis has been performed to assess the impact of process variations and economic parameters on the PV.

The dilute-acid pretreatment process has the lowest PV among all process scenarios, which is estimated to be \$1.36/l of gasoline equivalent [LGE] (\$5.13/gal of gasoline equivalent [GGE]). Sensitivity analysis shows that the PV is most sensitive to feedstock cost, enzyme cost, and installed equipment costs. A significant fraction of capital costs is related to producing heat and power from lignin in the biomass.

Cellulosic ethanol production has yet to be commercialized. Hence, a pioneer plant is expected to be more costly to build and operate than an *n*th plant. To assess the impact of technological maturity on pioneer plant cost, a cost growth analysis was performed. The estimated value of PV for the pioneer plant is substantially larger than for the *n*th plant. The PV for the pioneer plant model with dilute-acid pretreatment is \$2.30/LGE (\$8.72/GGE) for the most probable scenario, and the estimated TCI was more than double the *n*th plant cost.

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Abbreviations: GGE, gal of gasoline equivalent; LGE, liter of gasoline equivalent; PV, product value with NPV = 0 in 20 years and 10% IRR; MACRS, modified accelerated cost recovery system; AFEX, ammonia fiber explosion; CAFI, consortium of applied fundamentals and innovation; FPU, filter paper units; TCI, total capital investment; CSL, corn steep liquor; CBP, consolidated bioprocessing; SHF, separate hydrolysis and fermentation; SSCF, simultaneous saccharification and co-fermentation; EVD, experimentally verified data; NPV, net present value; IRR, internal rate of return.

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1. Introduction

There is a growing interest in liquid biofuels produced from renewable biomass resources to reduce dependence on crude oil. The US demand for gasoline is about 103 billion gal (390 million m³) annually [1]; and approximately 9 billion gal (34 million m³) per year of ethanol is being produced from corn in 2008 [2]. The production of fuels from food crops may place upward pressure on the price and availability of food, and the likelihood of such a “food versus fuel” conflict may increase as the world’s population grows. As an alternative to corn grain, lignocellulosic biomass shows promise as a feedstock for bioethanol production. It is projected that by 2030, over 38 billion gal (144 million m³) per year of renewable biofuels will be consumed in the United

States, with less than half coming from conventional corn-based ethanol [1]. One study estimated that it is possible to grow enough lignocellulosic biomass – in an economically feasible and environmentally sustainable manner – to produce more than 50 billion gal (189 million m³) of biofuels annually [3].

Ethanol – along with other types of biofuels such as butanol, bio-gasoline, and dimethylfuran – can be derived from lignocellulose via different reaction pathways [4,5]. However, bioethanol research is more advanced than many competing biofuel technologies, most of which are at the early stages of development. Ethanol can be produced from lignocellulosics following two different process pathways: (i) biochemical, where chemical or enzymatic hydrolysis and subsequent microbial fermentation is applied; and (ii) thermochemical, where gasification followed either by microbial fermentation or by catalytic upgrading are applied [6]. Unlike ethanol production from starch feedstocks, lignocellulosic biomass requires more aggressive pretreatments prior to saccharification and fermentation to increase the exposure of cellulose to enzymes during enzymatic hydrolysis. Dissolved catalysts are often used during biomass pretreatment. The effectiveness of the catalyst and pretreatment conditions contributes significantly to the yield and economics of the overall process. Both acids and bases are used as catalysts, with acids resulting in significantly different product yields than bases. Sulfuric acid, sulfur dioxide, ammonia, and lime are some of the catalysts that have been studied [7]. Hot water pretreatment [8], which relies on the reduced pH of water at elevated temperatures to hydrolyze hemicelluloses and disrupt the biomass structure, can also be used.

Following pretreatment, the biomass cellulose and hemicellulose components are hydrolyzed to monosaccharides (primarily glucose and xylose) either by acids or enzymes. The sugars are then fermented by yeast or by bacteria to produce ethanol. With separate fermentation of C5 and C6 (5-carbon and 6-carbon) sugars using selective microbes, higher ethanol yields can be achieved than with co-fermentation [9,10]. The produced ethanol is purified in distillation columns; and advanced purification technologies such as pervaporation [11,12], and reverse osmosis [13–15] are being developed, which may reduce operating costs.

This study, which is a techno-economic analysis of biochemical ethanol production from corn stover, focuses on technologies projected to be viable within a 5- to 8-year time frame. Based on this time frame – and after considering time for design, construction, and start-up – the process would likely have to be based on experimental data available today. Initially, 35 published technologies of various liquid fuels were reviewed, and a matrix was prepared considering economics, technological maturity, environmental aspects, process performance, and technical and economic risks. Both butanol and ethanol production processes were initially included in the technology matrix. However, butanol technologies are at the lab-scale or very early pilot stage of development, and published data on butanol-producing organisms indicate low yields relative to ethanol production, so they are not included in further analysis. Seven lignocellulosic ethanol process scenarios were selected, with four involving pretreatment variations (di-

lute-acid, 2-stage dilute-acid, hot water, and ammonia fiber explosion or AFEX); and three involving downstream process variations (pervaporation, separate C5 and C6 fermentation, and on-site enzyme production). Fig. 1 shows a basic schematic of the cellulosic ethanol process with the model variations considered in this study listed below their respective process step.

Each of these scenarios is modeled in detail and economic analysis is performed assuming an *n*th plant design, meaning that the technologies used in the design have been employed in previous commercial plants and are relatively well understood. However, cellulosic ethanol production has yet to be commercialized, and a pioneer plant is expected to be significantly more expensive than an *n*th plant. To assess the impact of using immature technologies on the PV for a pioneer plant, the potential increase in capital cost and decreased plant performance were estimated using models developed by the RAND Corporation [16].

2. Materials and methods

A list of assumptions common to all process scenarios includes the following:

- Plant capacity is 2000 Mg/day of dry corn stover.
- 2007 Publicly available and experimentally validated reaction conversions and parameters are used.
- Equipment, chemical, and labor costs indexed to 2007 dollars.
- Process and steam generation plants depreciate in 7 and 20 years, respectively, following the modified accelerated cost recovery system (MACRS) method.
- Project is 100% equity financed.
- Contingency factor is 20% of total project investment.
- Capital investment is spread over 3 years at a rate of 8%, 60%, and 32% in the first, second, and third years, respectively.
- Working capital is 15% of fixed capital investment.
- Project life is 20 years.
- The internal rate of return is 10%.
- The average corn stover composition is based on stover from the Kramer farm in Wray, Colorado [17], at 25% moisture on a wet, as-received basis.

Process models were developed for the seven selected process scenarios using ASPEN Plus™ Process Simulator. Pinch analysis is used to optimize the process heat requirement, and the mass and energy flow rates from simulations are used to size process equipment. The costs of most of the equipment are obtained from vendor quotes from previous studies by NREL [18] and Consortium for Applied Fundamentals and Innovation (CAFI) [19,20]. Individual equipment costs are scaled based on the equipment size for the original price quote, using the scaling exponents appropriate for each type of equipment. The scaled equipment costs are indexed to 2007 dollars [21], and separate installation factors are used for each of the unit operations to obtain individual installed equipment costs [18].

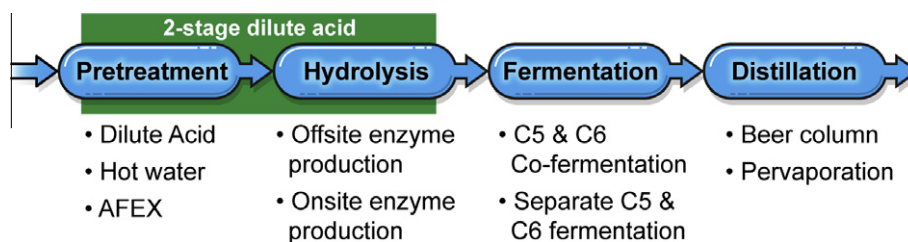


Fig. 1. Simplified block diagram for a cellulosic ethanol process plant.

The production costs include feedstock, variable operating costs (such as process chemicals, enzyme, nutrients, etc.), and fixed operating costs (employee salaries, overhead, maintenance, and insurance). The chemical and nutrient prices were obtained from previous NREL studies [18] and indexed to 2007 dollars [22]. Cellulase enzyme is not available commercially at scales needed for a production plant, which makes enzyme price information difficult to obtain. The enzyme price was estimated by using the on-site enzyme production model and increasing the biomass feed rate, so that the flow of pretreated biomass slurry to the saccharification vessels and the total ethanol production are the same as for the model without on-site enzyme production. This is done so that the amount of enzymes produced is equal to what is needed for a 2000 Mg/day plant without on-site enzyme production. The price of enzymes was then adjusted in the model for the 2000 Mg/day plant without on-site enzymes so that the PV is the same as the on-site enzyme model with adjusted biomass feed rate – that enzyme price is then used in all models. The raw material prices are shown in Table 1.

The PV (which includes production costs and a 10% rate of return) was calculated by iterating the ethanol selling price to achieve a net present value of zero. Because the lignocellulosic ethanol process has yet to be commercialized, a number of engineering design and performance uncertainties may arise. These uncertainties are accounted for in the cost growth analysis for a pioneer plant following a methodology developed by the RAND Corporation [16]. This methodology considers two sources of cost growth in chemical and mineral processing plants: a plant performance that is less than expected and a capital cost estimation that is low. Cost growth sources are estimated by using two multi-factor linear correlations to estimate the unexpected reduced plant performance and capital cost growth associated with pioneer plants.

2.1. Process description

The base case cellulosic ethanol process, modified from that used for the NREL 2002 design report [18], is comprised of nine sections: feed handling, pretreatment and detoxification, enzymatic hydrolysis and fermentation, on-site enzyme production, product recovery, wastewater treatment, product and raw material storage, burner/boiler turbogenerator, and utilities.

All seven process scenarios were simulated using current published data, which will be referred to as 2007 experimentally verified data (EVD). The reaction conversions for pretreatment and enzymatic hydrolysis for all scenarios (except for 2-stage dilute-acid treatment) are obtained from CAFI research [7,23]. For other downstream processes in the base case scenario, the data were obtained from NREL research [24]. In addition, the techno-economic

models used in a previous CAFI study [19,20] have been updated to reflect the latest NREL design report [18] and updated financial assumptions mentioned earlier.

For each of the process scenarios (except the on-site enzyme production scenario), the analysis assumed that enzymes are purchased from an external supplier and so, this section is absent. Pretreatment, enzymatic hydrolysis, on-site enzyme production and product recovery are subjected to process variations as mentioned in the methodology. The remaining sections are common for each of the seven process variations and are the same as the 2002 NREL design report [18]. For the pretreatment scenarios, it is possible that the optimal configuration is not represented by the simple substitution of a different process into the pretreatment section of the 2002 NREL design report. However, the modeling methodology taken in this study mirrors the experimental procedure and data from CAFI research.

The pretreatment area in the dilute-acid pretreatment scenario is modeled identically to the 2002 NREL design report [18], with the exception of different reaction conversions. In the 2-stage dilute-acid pretreatment scenario, the first stage solubilizes most of the hemicellulose just as in the base case pretreatment scenario. In the second stage, dilute-acid hydrolyzes a fraction of the cellulose and remaining hemicellulose. This contrasts with the other process scenarios in which enzymes are used to hydrolyze the pretreated cellulose. The 2-stage dilute-acid process was derived from a previous NREL report with softwood as the feedstock [25]. Because of the lack of publicly available experimental data with corn stover, conversions were based on experiments on softwood [26] and assumed that the conversions would also apply for corn stover. Uncertainty analysis was used to help capture differences in conversions as a result of the feedstock.

In the hot water pretreatment scenario, the chopped and washed biomass from the pretreatment section is mixed with recycled hot water from the distillation column bottoms [8]. The slurry is fed to a plug flow pretreatment reactor, where the pressure is maintained at 12.5 bars and the temperature is held constant at 190 °C; the residence time in the pretreatment reactor is 5 min.

In the AFEX pretreatment scenario, the biomass is treated with liquid ammonia under high pressure (17.2 bars) at 60 °C for 5 min [27]. The pressure is rapidly released causing the fibers to explode, which increases the accessibility of the cellulase enzymes to cellulose. Most of the ammonia is recovered from the blow down tank. Residual ammonia is recovered from the solids in a flash drum followed by fractionation of other volatiles. Recovered ammonia vapor is compressed, condensed, and recycled back into AFEX reactor [27].

The hydrolyzate from dilute-acid, hot water, and AFEX pretreatments is pumped into one of several parallel saccharification vessels where enzymes are added in the saccharification and fermentation section of the plant [18]. The enzyme loading is 31.3 mg protein/g cellulose in the untreated biomass; the residence time for saccharification is 5 days. Although enzymes could be tailored and optimized for each pretreatment option, this was not done experimentally in the CAFI work [19,20,23] and the resulting effect on enzyme cost would be difficult to characterize. In the absence of additional data, assuming the same enzyme and loading permits the best comparison of pretreatment options.

Recombinant *Zymomonas mobilis* is capable of fermenting glucose and xylose to ethanol. The hydrolyzate from the saccharification vessels is pumped into one of several parallel-sequenced batch fermentation reactors, where *Z. mobilis* ferments xylose and glucose to ethanol with conversions of 0.756 and 0.95, respectively [18].

A scenario with a fermentation process variation was explored, with xylose and glucose being fermented separately using the selective fermentative microbes *Z. mobilis* for xylose and yeast

Table 1
Feedstock and chemical costs.

	Price
Corn stover (US\$/Mg)	83
Enzyme broth (US\$/Mg) at 100 g protein/L broth	507
Sulfuric acid (US\$/Mg)	35
Hydrated lime (US\$/Mg)	99
Corn steep liquor (CSL) (US\$/Mg)	226
Diammonium phosphate (US\$/Mg)	201
Electricity price (c/kWh)	5.4
Propane (US\$/Mg)	374
Boiler chemicals (US\$/Mg)	4996
Cooling tower chemicals (US\$/Mg)	2994
Wastewater chemicals (US\$/Mg)	463
Wastewater polymer (US\$/Mg)	7485
Clarifier polymer (US\$/Mg)	3575

(*Saccharomyces cerevisiae* or *pastorianus*) for glucose [28]. This scenario is intended to avoid the issue of lower ethanol yields of recombinant *Z. mobilis* when fermenting both C5 and C6 sugars. However, the disadvantage of this scenario is that more water is needed to dilute the solids stream (C6 sugar) because the best yields are achieved at low solids loading. The additional water results in a lower ethanol concentration in the beer, which increases distillation costs. To increase the ethanol concentration in the beer, a fraction of the product stream from xylose fermentation is recycled to saccharification; however, this is done at the expense of lower cellulose-to-ethanol yields.

As an alternative to purchasing enzymes, the production of enzymes on-site was studied as a process variation. It is thought that production of enzymes on-site might reduce enzyme cost by eliminating transportation and the need to add stabilizers to reduce enzyme degradation during storage. In this scenario, *Trichoderma reesei* is used for on-site enzyme production, which is modeled the same as the 1999 NREL design report [29]. For inoculum growth and enzyme production, a fraction of conditioned pretreated biomass is used as a carbon source. The specific activity of the enzymes, enzyme yield, and productivity are 600 filter paper units (FPU)/g protein, 0.33 g protein/(g cellulose and xylose), 0.125 g protein/(L h), respectively [29].

Ethanol is recovered from beer and dissolved solids by two-step distillation: a beer splitter column followed by a rectification column. Distillation is an energy-intensive operation. To minimize energy consumption in this section, a pervaporation separation scenario to replace the beer column is evaluated. Pervaporation refers to separation using a membrane with liquid feed on one side, and a low-pressure, gaseous permeate output on the other. Components in the liquid feed preferentially permeate through the membrane, and then evaporate into the gaseous phase. Pervaporation does not involve large heat input, which reduces costs associated with the heat and steam needed for reboilers in the base case distillation columns. In the pervaporation scenario, a membrane is inserted in place of the beer column. The pervaporation system output is calculated from separation and total material flux factors to achieve the same separation as the existing beer column of 40 wt.% ethanol [30]. The resulting separation factor and total material flux were reasonable, given values used in a previous economic study [31]. In the base case scenario, the beer column also separated out carbon dioxide for scrubbing. However, in the pervaporation scenario, a flash tank is added to separate out CO₂ at 110 °C and a heat exchanger cools the stream to 41 °C. The membrane system costs \$200/m² in 1999 dollars, with a replacement needed every 5 years at a cost of \$100/m² [31].

Wastewater is treated and recycled as process water, while steam and electricity are produced using the remaining biomass components as boiler fuel.

2.2. Pioneer plant analysis

RAND Corporation developed two multi-factor linear regressions to estimate production shortfalls and capital growth of pioneer process plants [16]. The regressions were developed using data collected from 44 process plants. The production shortfalls are accounted for in the regression equation for “plant performance”, which is the production in the second six months after initial operation as a percentage of design capacity. Parameters used to estimate plant performance are based on the number of process steps that have not been demonstrated commercially, the mass and energy balance equations that are validated with commercial scale data, the potential for waste-handling issues, and whether the plant handles solids.

The second regression equation (“cost growth”) estimates the increase in actual capital cost over the original estimate. Param-

eters for cost growth are based on the cost estimate for equipment not demonstrated at the commercial scale, the potential for buildup of impurities, the number of continuously linked steps, and the inclusiveness of pre-start-up and land costs, and the level of site-specific information that is included in the cost estimate.

Three scenarios are considered in the pioneer plant analysis – most probable, optimistic, and pessimistic – representing the range of estimates for variables used in the regression equations for cost growth and plant performance. However, only a comparison of the *n*th plant numbers and the most probable case is discussed in the paper.

There are two variables that have the most significant impact on plant performance. The first variable accounts for the number of steps not demonstrated on a commercial scale. The new steps for the process and units considered here are feedstock handling, pretreatment, saccharification, co-fermentation, beer column, and the fluidized bed combustor. The second variable accounts for the mass and energy balances that can be verified with commercial production data. This variable was assigned a low value for the most probable scenario because, with the exception of distillation, none of the mass and energy flows can be commercially verified.

For the equation for cost growth, there are two variables that are the most significant. One of the variables accounts for the percentage of total cost of new technologies employed in the plant. Feedstock handling, pretreatment, the saccharification vessels, co-fermentation, beer column and the fluidized bed combustor are included as new technologies. The variable accounting for potential buildup of impurities that can affect the process was given a mid-range value. There is a possibility of degradation products such as furfurals building up in the process, which are inhibitory to fermentative organisms. The other parameters for cost growth were less impactful for the range of variables chosen.

In the discounted cash flow spreadsheet, the total capital investment (TCI) of the base case *n*th plant is divided by the percentage of cost growth to estimate the TCI of the pioneer plant. The first-year ethanol sales, variable operating costs, and electricity export of the *n*th plant is multiplied by the percentage of plant performance to account for the reduced production of a pioneer plant. For the discounted cash flow analysis, the plant performance is increased by 20% per year until design capacity is reached. The contingency is increased to 30% from 20% used in the *n*th plant analysis to account for greater uncertainty in equipment and other costs related solely to the *n*th plant design, even before considering cost growth or plant performance.

3. Results and discussion

Results of the techno-economic process model and product values (PV) are presented in Table 2 and Fig. 2, respectively.

There is significant variation of ethanol yields per mass of dry feedstock among the pretreatment processes, with 2-stage dilute-acid pretreatment being the lowest (47 gal/Mg or 0.18 m³/Mg) and dilute-acid pretreatment (base case scenario) being the highest (76 gal/Mg or 0.29 m³/Mg). The installed equipment cost of the dilute-acid pretreatment scenario is \$164 million, and the costs for other pretreatment scenarios vary between \$156 and \$173 million with the hot water pretreatment scenario being the lowest. The most significant contributor to the lower equipment cost of the hot water pretreatment scenario is relative simplicity, and therefore lower cost, of the horizontal tubular reactor. The installed cost of the tubular reactor is \$0.31 million compared with the dilute-acid (base case) and AFEX pretreatment reactor costs of \$22.99 and \$9.15 million, respectively. Although the AFEX reactor cost is lower than the dilute-acid pretreatment reactor, the additional expense of the ammonia recycle equipment results in

Table 2
Techno-economic analysis results for all processes variations.

Process variations	Total capital investment (\$ million)	Total installed equipment cost (\$ million)	Ethanol yield (l/Mg)	Ethanol production (million l/year)	Electricity export (\$ million/year)
Dilute-acid pretreatment (base case)	376	164	289	202	11.7
Dilute-acid pretreatment (high solids) ^a	389	169	274	192	12.6
2-Stage dilute-acid pretreatment	391	173	177	124	16.8
Hot water pretreatment	327	156	211	148	11.3
AFEX pretreatment	386	167	250	175	16.9
Pervaporation–distillation	501	209	291	204	13.6
Separate C5 and C6 fermentation	386	168	300	210	6.5
On-site enzyme production	434	188	256	179	–0.8

^a Pretreatment conversions used in the model are from NREL's pretreatment reactor.

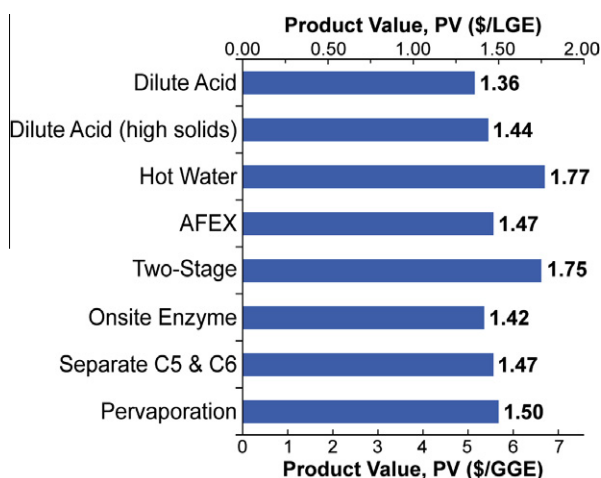


Fig. 2. Estimated ethanol product values (PV) from various process scenarios.

a total installed equipment cost that is greater than that of the dilute-acid pretreatment scenario.

The PV for the dilute-acid pretreatment scenario is \$1.36/LGE, which is the lowest among all scenarios (Fig. 2). It was thought that the 2-stage dilute-acid pretreatment scenario may offer economic advantages because the process does not use enzymes for glucose hydrolysis, which is estimated to represent one of the most significant expenses. Additionally, capital costs in the pretreatment and hydrolysis stages are reduced when \$13.2 million of installed enzymatic hydrolysis equipment are replaced with \$8.4 million of 2-stage acid hydrolysis equipment. However, the lower cellulose yields in 2-stage acid hydrolysis reduce the ethanol production capacity, thereby offsetting the cost reductions due to not using enzymes. The ethanol yield with the 2-stage dilute-acid scenario is only 177 l/Mg compared to 289 l/Mg for the dilute-acid base case scenario, while the installed equipment cost is higher than that of the dilute-acid base case scenario.

The capital and operating cost breakdown by process areas and cost components for the dilute-acid pretreatment scenario is shown in Figs. 3 and 4, respectively. The most expensive sections are the boiler/turbogenerator and pretreatment. Additional capital expense is incurred for the dilute-acid pretreatment scenario because of the need to condition the pretreated slurry prior to fermentation. Overliming to remove sulfuric acid from the slurry adds an additional \$10.8 million to the cost of dilute-acid pretreatment. The electricity credit from the boiler/turbogenerator for dilute-acid pretreatment scenario is only 7%, whereas feedstock and variable operating cost is 73% of the total operating cost (Fig. 4).

It should be noted that the installed cost of the boiler/turbogenerator is 34.2% of the total. An often-cited advantage of cellulosic

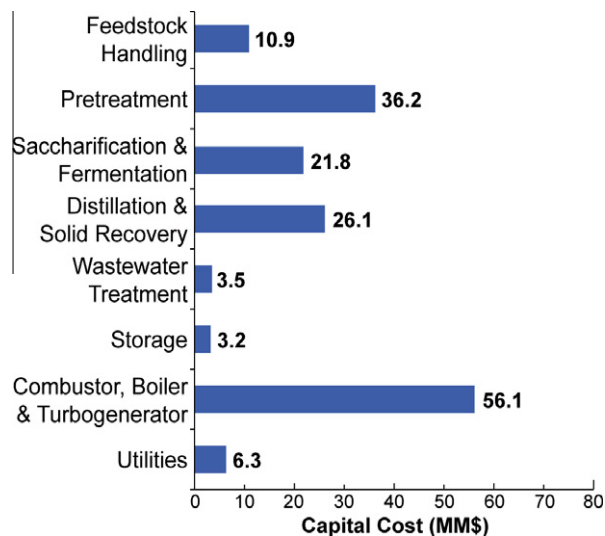


Fig. 3. Capital cost breakdown for the dilute-acid pretreatment scenario.

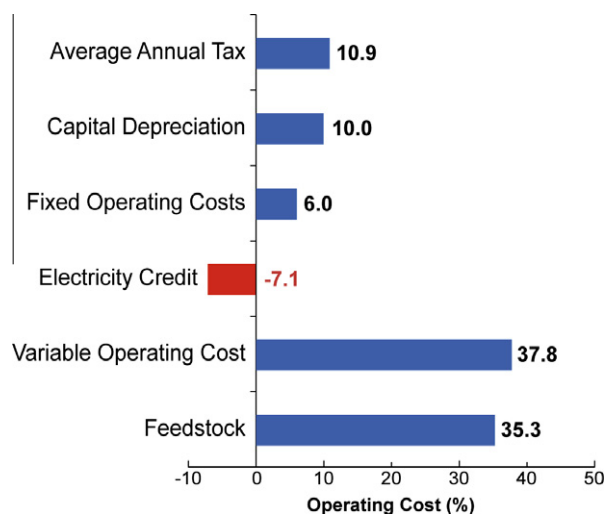


Fig. 4. Operating cost component as percentage of total operating cost for dilute-acid pretreatment scenario.

ethanol over corn grain ethanol is that the internally used energy is supplied by the plant by-products. However, this advantage in renewable energy to fossil fuel input ratio may prove to be a major financial impediment to commercialization because of the significant capital costs. An alternative to burning plant by-products for heat and power is burning natural gas. A natural gas boiler was

costed at \$3 million installed [32], compared to over \$30 million for a fluidized bed combustor needed for solid fuel. At a natural gas cost of \$7.05/BTU (based on 2007 prices [33]), the PV would increase to \$1.43/LGE from \$1.36/LGE, even before accounting for disposal costs of the lignin and other plant by-products. A natural gas boiler, as an understood commercial process, may have a cost advantage in the pioneer plant. Those cost advantages would be diminished by the need to deal with fluctuations in natural gas price and a more greenhouse gas intensive process that may not qualify as a renewable fuel. Ultimately, the heat and power requirements for the cellulosic ethanol process may best be addressed by lignin and other plant by-products, albeit with substantial capital equipment costs that affect PV.

Alternative saccharification/fermentation and distillation scenarios were explored to determine their impact on PV. A trade-off exists with separate C5 and C6 sugar fermentation between higher ethanol yields, and increased capital cost of additional fermentation vessels, as well as the additional operating expenses resulting from lower ethanol concentration in the beer (4–7%). Ethanol yield in the base case scenario was 289 l/Mg. And while separate C5 and C6 fermentation using selective microbes increased ethanol yield to 300 l/Mg, the PV is \$0.11/LGE higher than the base case scenario.

Distillation is a cost-intensive operation, and the cost increases as ethanol concentration in the beer decreases. As a less energy-intensive process, pervaporation could decrease the operating costs compared to distillation. In this scenario, the pervaporation membrane is used in place of the beer column. Because of the reduced energy consumption of pervaporation, more steam is available for electricity generation, and the exported electricity value increases nearly \$2 million/year from the base case. However, the pervaporation option increased the PV by \$0.14/LGE over the base case. This is a result of the high capital cost of the membrane. The installed membrane cost is \$46.5 million compared to the beer column cost of only \$1.5 million.

It is assumed that for the base case scenario, enzyme broth is purchased from external sources. However, production of enzymes on-site may offer economic advantages because it eliminates the

cost of broth concentration, enzyme stabilizers, and transportation – costs that are not included this analysis. The annual enzyme cost for the base case scenario is \$0.28/LGE. The PV from the on-site enzyme production process scenario is \$1.42/LGE, which is \$0.06/LGE higher than the base case scenario. This difference is a result of a significant fraction of the feedstock (9.2% of the hydrolyzate) being diverted to the enzyme production area, which reduces the plant capacity by 22.7 million l/year of ethanol. The reduction in ethanol capacity reduces economy-of-scale advantages that occur with increasing plant size. This could be overcome by increasing the biomass feed so that the ethanol capacity is equivalent to the base case scenario. This was done to calculate the enzyme cost for the base case scenario, and thus led to a lower PV. However, a greater biomass feed rate requires greater biomass availability. The cost of the enzyme is also affected by a lower electricity credit than the base case, which is due to high electricity consumption by the compressor supplying air to the enzyme production bioreactors that leads to lower net excess electricity and lower ethanol yield. Although this comparison does not include additional costs associated with purchased enzymes, the comparison to an on-site case helps demonstrate what yield and electricity tradeoffs occur when enzyme is produced on-site.

3.1. Uncertainty analysis

Process-specific uncertainty analysis – involving pretreatment and saccharification process conditions and reaction conversions – has been performed on all pretreatment process scenarios to study the impact of operating temperature, retention time, acid concentrations, and yields on PV. These uncertainties are based on a range of results found in the literature [7,8,25,26], which are shown in Figs. 5 and 6.

The PV is most sensitive to pretreatment retention time, xylan conversions, solids loading, and cellulose conversions. When the residence time of the dilute-acid pretreatment reactor is increased from 2 to 10 min, an increase in PV of 15% is observed because larger reactors are needed. When the conversion of xylan to xylose in the pretreatment reactor is reduced from 82.5 (2007 EVD) to 33%,

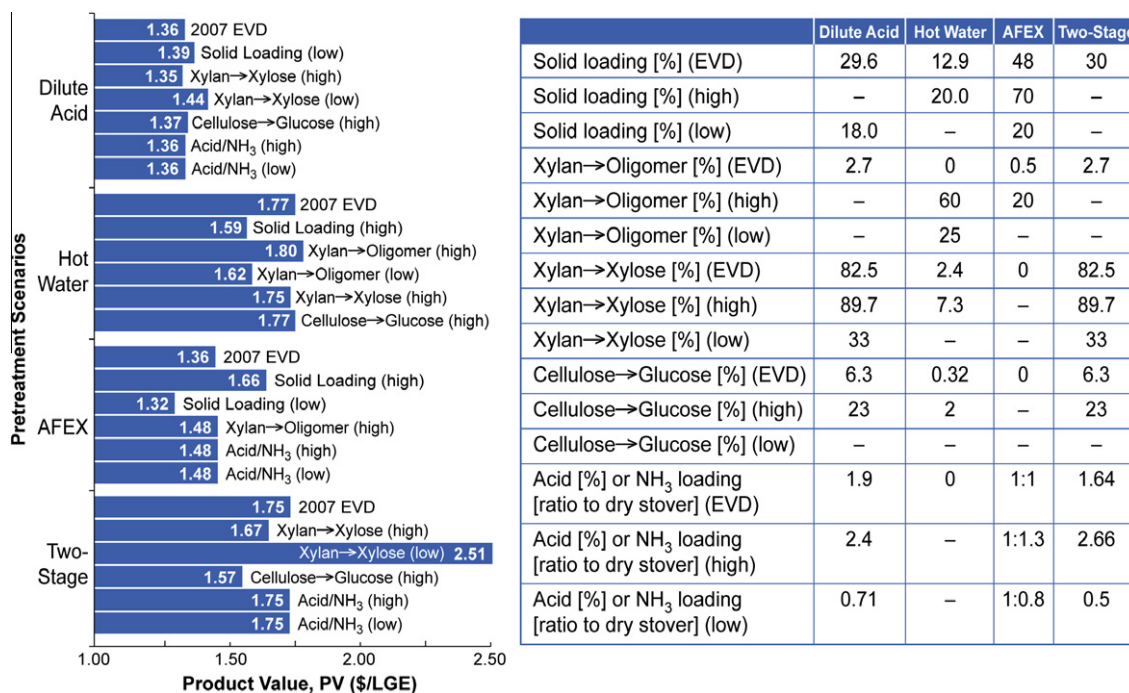


Fig. 5. Impact of pretreatment parameters on ethanol product value (PV) for different pretreatment scenarios.

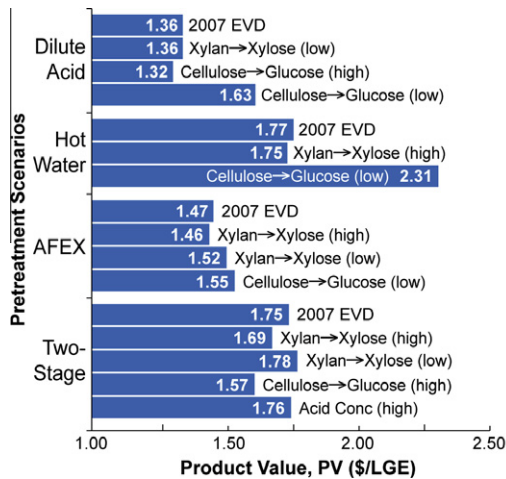


Fig. 6. Impact of saccharification parameters on ethanol product value (PV) for different pretreatment scenarios.

the PV increases from \$1.36/LGE to \$1.44/LGE (a 6% increase). Increased solids consistency during pretreatment resulted in lower PV because a smaller reactor volume is necessary and the requirement for process heat is lower. The impact of other pretreatment reactor parameters on PV is not very significant. Among the saccharification reactor parameters, cellulose-to-glucose conversions showed significant impact on PV for all scenarios (Fig. 6).

An overall economic uncertainty analysis on the dilute-acid pretreatment scenario has also been performed. The selected uncertainty parameters were feedstock cost, enzyme loading, enzyme cost, contingency factor, installed pretreatment reactor cost, total installed equipment cost, and export electricity price (the results are presented in Fig. 7). The feedstock and enzyme costs have the most significant impact on PV. When feedstock cost increased from \$83/Mg (base case scenario) to \$110/Mg (\$100/dry short ton), PV increased by 11%. The PV reduced the same percentage when the feedstock cost was \$55/Mg. When enzyme cost was reduced from \$507/Mg (\$460/short ton) to \$256/Mg (equivalent to \$0.14/LGE

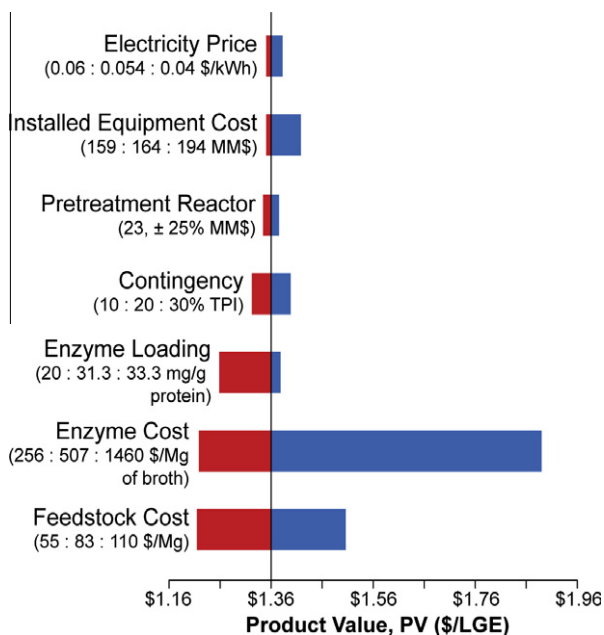


Fig. 7. Impact of overall process/economic parameters on ethanol product value (PV) for dilute-acid pretreatment scenario.

	Dilute Acid	Hot Water	AFEX	Two-Stage
Xylan→Xylose [%] (EVD)	57.13	56.61	77.7	14.3
Xylan→Xylose [%] (high)	–	63	90	41.5
Xylan→Xylose [%] (low)	52.4	–	55	0
Cellulose→Glucose [%] (EVD)	91.09	89.97	95.9	42.4
Cellulose→Glucose [%] (high)	97	–	–	48
Cellulose→Glucose [%] (low)	67	65	75	–
Acid [%] or NH ₃ loading [ratio to dry stover] (EVD)	–	–	–	1.80
Acid [%] or NH ₃ loading [ratio to dry stover] (high)	–	–	–	2.6
Acid [%] or NH ₃ loading [ratio to dry stover] (low)	–	–	–	–

Table 3

PV, capital costs, and Lang factor obtained from cost growth analysis for pioneer plant using dilute-acid pretreatment.

Cost item	Pioneer plant	<i>n</i> th Plant
PV (\$/LGE)	2.30	1.36
Fixed capital investment (\$ millions)	833	327
Total capital investment (\$ millions)	886	376
Lang factor	8.11	3.44

produced) PV decreased by 10%; and when enzyme cost was \$1460/Mg (equivalent to \$0.83/LGE produced) the PV increased by 39%. The price of cellulase and xylanase cocktails produced for large-scale cellulosic ethanol plants is not yet known, and the wide range of prices used represents publicly available estimates.

The installed equipment cost was varied to represent different installation factors. When the installed equipment cost increased from \$164 million (corresponding weighted average installation factor of 1.5 for base case scenario) to \$194 million (corresponding installation factor of 2.05 [34]), the PV increased 6%. When the contingency factor was varied between 10% and 30%, the PV changed by –5–8%, respectively. The analysis showed less significance from the impact of other parameters.

3.2. Cost growth analysis

Table 3 shows the results of the pioneer plant analysis for the dilute-acid pretreatment scenario.

The PV for the most probable scenario is \$2.30/LGE, which is 69% more than the PV estimated for the *n*th plant cost analysis. For the most probable case pioneer plant, the TCI was 136% greater than the *n*th plant dilute-acid pretreatment scenario. The Lang factor (the ratio of total capital investment to purchased equipment costs) for the pioneer plant also increased from 3.44 to 8.11 denoting the relative uncertainty involved in the construction of a pioneer plant as opposed to an *n*th plant.

4. Comparison with previous studies

The results of this study deviate considerably from a number of previous techno-economic analyses of cellulosic ethanol production. There are many contributing factors to this deviation and an explanation of the most significant of these factors is discussed here. Fig. 8 presents a plot of estimated ethanol prices from seven previous studies as a function of feedstock price. The ethanol and

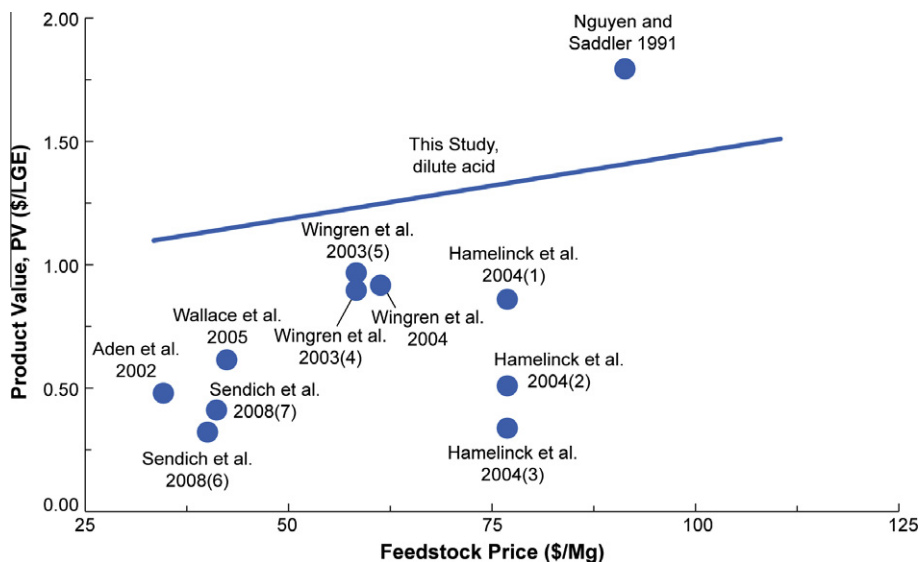


Fig. 8. Ethanol product values (PVs) from previously published techno-economic studies and comparison. *Note:* (1) short-term technology – simultaneous saccharification and fermentation (SSF) (2) middle-term technology – simultaneous saccharification and co-fermentation (SSCF) (3) long-term technology – consolidated bioprocessing (CBP) (4) separate hydrolysis and fermentation (SHF) (5) SSF (6) CBP and (7) SSCF.

feedstock prices were updated to 2007 dollars using the Consumer Price Index. The solid line on the plot represents the PV for the dilute-acid pretreatment scenario using the model developed in this study as a function of feedstock price.

After updating the feedstock and ethanol prices to 2007 dollars, much of the difference from previous studies can be explained by the clear correlation that exists between feedstock price and ethanol price. However, all of the studies except that of Nguyen and Saddler [35] remain lower than the line derived from this study. The study by Hamelinck et al. [36] represents a significant outlier from the apparent correlation between feedstock price and ethanol price. The three ethanol price estimates are for short- (5 years from time of study), middle- (10–15 years), and long-term (20+ years) technology implementation. The short-term estimate is closer to the time frame considered in this study. However, it also deviates from the trend of other studies. The assumptions for the short-term estimate – including feedstock input, rate of return, and reaction conversions – are quite similar to those in this study; and the TCI (updated to 2007 dollars) is nearly equal as well. The most significant difference from this study is the non-feedstock operating cost, which is approximately \$0.13/LGE compared to \$0.67/LGE. This is partly due to lower costs for corn steep liquor, cellulase, and other raw materials. This factor accounts for most of the discrepancy between ethanol price estimates.

The ethanol price from the study published by Sendich et al. [37] is also slightly lower than the apparent correlation of feedstock and ethanol price. The lowest estimate in that study assumes the use of consolidated bioprocessing, which is an advanced technology also modeled in the long-term estimate from Hamelinck et al. [36]. The higher ethanol price estimate of \$0.41/LGE is from a model using simultaneous saccharification and co-fermentation (SSCF). SSCF is also a more advanced technology than was considered in this study; this factor results in lower capital and operating costs by combining enzymatic saccharification and fermentation. A new AFEX pretreatment scheme was also employed, which may have contributed to lower capital and operating costs of pretreatment.

The enzyme cost used in this study is much higher than that used in other studies; and because enzyme cost is such a significant fraction of the PV, it contributes significantly to the discrepancy between the current study and previous studies. For

example, the enzyme prices used in prior studies [38,39,18] are approximately 30%, 30%, and 17% of the price used in this study, respectively.

5. Conclusions

The PV for the dilute-acid pretreatment scenario is \$1.36/LGE, which is the lowest among all pretreatments and process variations. This is due primarily to the higher sugar yields – and, therefore, ethanol yields – from dilute-acid pretreatment and enzymatic hydrolysis than for the other process scenarios. The exception to this is the scenario with separate C5 and C6 sugar fermentation, which has higher ethanol yields. However, the PV is higher because of the high capital cost of extra fermentation vessels needed to ferment the sugars separately.

A high level of uncertainty exists for the cost of cellulase enzymes, which leads to a wide range of PV observed in the sensitivity analysis. The range of enzyme price used in the sensitivity analysis corresponds to enzyme costs of \$0.14–\$0.80/LGE. The range of PV following from these enzyme prices is \$1.22–\$1.89/LGE. Similarly, there is continued debate regarding the price of corn stover. The range of PV from the sensitivity of the feedstock price is \$1.22–\$1.51/LGE. While these raw materials contribute to a high degree of financial uncertainty of cellulosic ethanol production, it may prove difficult to fully understand their costs until markets exist.

Installed equipment costs for an *n*th plant total \$164 million. The largest contributor to this cost at \$56.1 million is the boiler/turbogenerator system needed to convert lignin to heat and power for the process. Although producing electricity from plant by-products provides sustainability advantages and may be the most cost effective process, its large capital cost may hinder commercialization.

To estimate the potential risk associated with process scale-up, a pioneer plant risk analysis was conducted. Under the most probable assumptions for pioneer plant operation for the dilute-acid pretreatment scenario, the PV is \$2.30/LGE. Additionally, the TCI for the most probable case is estimated to double from the cost of an *n*th plant. Because of the large capital cost and the PV being well above ethanol market prices for a pioneer plant, it may prove difficult for the cellulosic ethanol industry to finance growth until a number of biotechnology barriers are broken.

Significant opportunities exist to reduce the PV through biochemical technology breakthroughs. Enzyme cost in this study is assumed to be \$0.27/LGE, representing a potential ethanol cost reduction by reducing enzyme production cost and increasing specific activity. In the dilute-acid pretreatment model in this study, 75.6% of xylose is converted to ethanol during fermentation and none of the other hemicellulose sugars are converted to ethanol. The development of organisms that can ferment xylose at conversions similar to those of glucose to ethanol— as well as other hemicellulose sugars – also offers potential for reducing ethanol cost.

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References

- Holtberg PD, Smith KA, Mayne L, Doman L, Cymbalsky JH, Boedecker EE, et al. Annual energy outlook with projection to 2030. Energy Information Administration. DOE/EIA-0383; 2009.
- Renewable Fuels Association, Growing innovation: 2009 ethanol industry outlook; 2009.
- Perlack RD, Wright LL, Turhollow AF, Graham RL, Stokes BJ, Erblich DC. Biomass as feedstock for a bioenergy and bioproducts industry: the technical feasibility of a billion-ton annual supply. US Department of Energy, US Department of Agriculture. DOE/GO-102005-2135 ORNL/TM-2005/66; 2005.
- Ezeji TC, Qureshi N, Blaschek HP. Acetone–butanol–ethanol (ABE) production from concentrated substrate: reduction in substrate inhibition by fed-batch technique and product inhibition by gas stripping. *Appl Microbiol Biotechnol* 2004;63:653–8.
- Huber GW, Chheda JN, Barrett CJ, Dumesic JA. Production of liquid alkanes by aqueous phase processing of biomass-derived carbohydrates. *Science* 2005;308:1446.
- Brown RC. In: Kamm B, Gruber PR, Kamm M, editors. Biomass refineries based on hybrid thermochemical/biological processing – an overview, in biorefineries, biobased industrial processes and products. Weinheim, Germany: Wiley-VCH Verlag; 2005.
- Lloyd TA, Wyman CE. Combined sugar yields for dilute sulfuric acid pretreatment of corn stover followed by enzymatic hydrolysis of the remaining solids. *Bioresour Technol* 2005;96:1967–77.
- Moiser N, Hendrickson R, Ho N, Sedlak M, Ladisch M. Optimization of pH controlled hot water pretreatment of corn stover. *Bioresour Technol* 2005;96:1986–93.
- Kadar Z, Maltha SF, Szengyel Z, Reczey K, Laar W. Ethanol fermentation of various pretreated and hydrolyzed substrates at low initial pH. *Appl Biochem Biotechnol* 2007;136–140:847–58.
- Olsson L, Hagerdal BH. Fermentation of lignocellulosic hydrolysates for ethanol production. *Enzyme Microb Technol* 1996;18:312–31.
- Kalyani S, Smitha B, Sridhar S, Krishnaiah A. Pervaporation separation of ethanol–water mixtures through sodium alginate membranes. *Desalination* 2008;229:68–81.
- Lee YM, Shin EM, Yang KS. Pervaporation separation of water–ethanol through modified chitosan membrane: 1. Chitosan–acetic acid and metal ion complex membranes. *Polymer (Korea)* 1991;15:182–90.
- Ling KC. Whey to ethanol: a biofuel role for dairy cooperatives? Research report 214. Agricultural Economist. USDA rural development; 2008.
- Hsu J-H, Tan C-S. Separation of ethanol from aqueous solution by a method incorporating supercritical CO₂ with reverse osmosis. *J Membr Sci* 1993;81:273–85.
- Choudhury JP, Ghosh P, Guha BK. Ethanol separation from molasses based fermentation broth by reverse osmosis. *Biotechnol Lett* 1986;8:731–4.
- Merrow WE, Phillips KE, Myers CW. Understanding cost growth and performance shortfalls in pioneer process plants. USA: The Rand Corporation; 1981.
- Weiss ND, Nagle NJ, Tucker MP, Elander RT. High xylose yields from dilute acid pretreatment of corn stover under process-relevant conditions. *Appl Biochem Biotechnol Part A: Enzyme Eng Biotechnol* 2009;155:115–25.
- Aden A, Ruth M, Ibsen K, Jechura J, Neeves K, Sheehan J, et al. Lignocellulosic biomass to ethanol process design and economics utilizing co-current dilute acid prehydrolysis and enzymatic hydrolysis for corn stover. Golden, Colorado: National Renewable Energy Laboratory; 2002.
- Eggeman T, Elander RT. Process and economic analysis of pretreatment technologies. *Bioresour Technol* 2005;96:2019–25.
- Elander RT, Eggeman T. Alternative pretreatment (CAFI)-initial data and economic analysis. Golden, Colorado: National Renewable Energy Laboratory; 2004.
- CEPCI (Chemical Engineering Plant Cost Index). Chemical engineering. November issue; 2008.
- ICI (Inorganic Chemical Index). SRI international economics handbook. Economic environment of the chemical industry; 2008.
- Wyman CE, Dale BE, Elander RT, Holtzapfle M, Ladisch MR, Lee YY. Comparative sugar recovery data from laboratory scale application of leading pretreatment technologies to corn stover. *Bioresour Technol* 2005;96:2026–32.
- Aden A. FY07 State of technology update. Golden, Colorado: National Renewable Energy Laboratory; 2007.
- Neeves K, Ibsen K. Lignocellulosic biomass-to-ethanol process design and economics using two-stage dilute acid hydrolysis. Technical report 29420. Golden, Colorado: National Renewable Energy Laboratory; 2000.
- Nguyen QA, Tucker MP, Keller FA, Eddy FP. Two-stage dilute-acid pretreatment of softwoods. *Appl Biochem Biotechnol* 2000;84–86:561–76.
- Teymouri F, Laureano-Perez L, Alizadeh H, Dale BE. Optimization of the ammonia fiber explosion (AFEX) treatment parameters for enzymatic hydrolysis of corn stover. *Bioresour Technol* 2005;96:2014–8.
- Dutta A, Dowe N, Ibsen K, Schell D, Aden A. An economic comparison of different fermentation configurations to convert corn stover to ethanol using *Z. mobilis* and *Saccharomyces* (p NA). *Biotechnol Prog* 2009. doi:10.1002/btpr.311.
- Wooley R, Ruth M, Sheehan J, Ibsen K, Majdeski H, Galvez A. Lignocellulosic biomass to ethanol process design and economics utilizing co-current dilute acid prehydrolysis and enzymatic hydrolysis current and futuristic scenarios. National Renewable Energy Laboratory Technical Report. NREL/TP-580-26157; 1999.
- Hsu D. Process economics of pervaporation membrane separation of ethanol and water in the biochemical design report. Golden, Colorado: Technical Memo, National Renewable Energy Laboratory; 2008.
- O'Brien DJ, Roth LH, McAloon AJ. Ethanol production by continuous fermentation–pervaporation: a preliminary economic analysis. *J Membr Sci* 2000;166:105–11.
- Aspen Icarus Project Evaluator 2006.5, Cambridge, MA: Aspen Technology, Inc.; 2007.
- US Energy Information Administration. US natural gas prices. <http://tonto.eia.doe.gov/dnav/ng/ng_pri_sum_nus_a.htm>. [accessed December 2009].
- Peters M, Timmerhaus K, West R, Peters M. Plant design economics for chemical engineers. 5th ed. New York: McGraw-Hill Professional; 2002.
- Nguyen QA, Saddler JN. An integrated model for the technical and economic evaluation of an enzymatic biomass conversion process. *Bioresour Technol* 1991;35:275–82.
- Hamelinck CN, van Hooijdonk G, Faaji APC. Ethanol from lignocellulosic biomass: techno-economic performance in short-, middle- and long-term. *Biomass Bioenergy* 2005;28:384–410.
- Sendich EN, Laser M, Kim S, Alizadeh H, Laureano-Perez L, Dale BE. Recent process improvements for the ammonia fiber expansion (AFEX) process and resulting reductions in minimum ethanol selling price. *Bioresour Technol* 2008;99:8429–35.
- Wingren A, Galbe M, Zacchi G. Techno-economic evaluation of producing ethanol from softwood: comparison of SSF and SHF and identification of bottlenecks. *Biotechnol Prog* 2003;19:1109–17.
- Wingren A, Soderstrom J, Galbe M, Zacchi G. Process considerations and economic evaluation of two-step steam pretreatment for production of fuel ethanol from softwood. *Biotechnol Prog* 2004;20:1421–9.